HANDBOOK OF ENVIRONMENTAL ENGINEERING™ VOLUME 9

Advanced Biological Treatment Processes

Edited by

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Humana Press

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ISBN: 978-1-58829-360-2 e-ISBN: 978-1-60327-170-7

DOI: 10.1007/978-1-60327-170-7

Library of Congress Control Number: 2008931192

Printed on acid-free paper

9 8 7 6 5 4 3 2 1

springer.com

Preface

The past 30 years have seen the emergence of a growing desire worldwide that positive actions be taken to restore and protect the environment from the degrading effects of all forms of pollution—air, water, soil, and noise. Because pollution is a direct or indirect consequence of waste, the seemingly idealistic demand for "zero discharge" can be construed as an unrealistic demand for zero waste. However, as long as waste continues to exist, we can only attempt to abate the subsequent pollution by converting it to a less noxious form. Three major questions usually arise when a particular type of pollution has been identified: (1) How serious is the pollution? (2) Is the technology to abate it available? and (3) Do the costs of abatement justify the degree of abatement achieved? This book is one of the volumes of the *Handbook of Environmental Engineering* series. The principal intention of this series is to help readers formulate answers to the last two questions above.

The traditional approach of applying tried-and-true solutions to specific pollution problems has been a major contributing factor to the success of environmental engineering, and has accounted in large measure for the establishment of a "methodology of pollution control." However, the realization of the ever-increasing complexity and interrelated nature of current environmental problems renders it imperative that intelligent planning of pollution abatement systems be undertaken. Prerequisite to such planning is an understanding of the performance, potential, and limitations of the various methods of pollution abatement available for environmental scientists and engineers. In this series of handbooks, we will review at a tutorial level a broad spectrum of engineering systems (processes, operations, and methods) currently being used, or of potential use, for pollution abatement. We believe that the unified interdisciplinary approach presented in these handbooks is a logical step in the evolution of environmental engineering.

Treatment of the various engineering systems presented will show how an engineering formulation of the subject flows naturally from the fundamental principles and theories of chemistry, microbiology, physics, and mathematics. This emphasis on fundamental science recognizes that engineering practice has in recent years become more firmly based on scientific principles rather than on its earlier dependency on empirical accumulation of facts. It is not intended, though, to neglect empiricism where such data lead quickly to the most economic design; certain engineering systems are not readily amenable to fundamental scientific analysis, and in these instances we have resorted to less science in favor of more art and empiricism.

Because an environmental engineer must understand science within the context of application, we first present the development of the scientific basis of a particular subject, followed by exposition of the pertinent design concepts and operations,

and detailed explanations of their applications to environmental quality control or remediation. Throughout the series, methods of practical design and calculation are illustrated by numerical examples. These examples clearly demonstrate how organized, analytical reasoning leads to the most direct and clear solutions. Wherever possible, pertinent cost data have been provided.

Our treatment of pollution-abatement engineering is offered in the belief that the trained engineer should more firmly understand fundamental principles, be more aware of the similarities and/or differences among many of the engineering systems, and exhibit greater flexibility and originality in the definition and innovative solution of environmental pollution problems. In short, the environmental engineer should, by conviction and practice, be more readily adaptable to change and progress.

Coverage of the unusually broad field of environmental engineering has demanded an expertise that could only be provided through multiple authorships. Each author (or group of authors) was permitted to employ, within reasonable limits, the customary personal style in organizing and presenting a particular subject area; consequently, it has been difficult to treat all subject material in a homogeneous manner. Moreover, owing to limitations of space, some of the authors' favored topics could not be treated in great detail, and many less important topics had to be merely mentioned or commented on briefly. All authors have provided an excellent list of references at the end of each chapter for the benefit of interested readers. As each chapter is meant to be self-contained, some mild repetition among the various texts was unavoidable. In each case, all omissions or repetitions are the responsibility of the editors and not the individual authors. With the current trend toward metrication, the question of using a consistent system of units has been a problem. Wherever possible, the authors have used the British system (fps) along with the metric equivalent (mks, cgs, or SIU) or vice versa. The editors sincerely hope that this duplicity of units' usage will prove to be useful rather than being disruptive to the readers.

The goals of the *Handbook of Environmental Engineering* series are: (1) to cover entire environmental fields, including air and noise pollution control, solid waste processing and resource recovery, physicochemical treatment processes, biological treatment processes, biosolids management, water resources, natural control processes, radioactive waste disposal, and thermal pollution control; and (2) to employ a multimedia approach to environmental pollution control because air, water, soil, and energy are all interrelated.

As can be seen from the above handbook coverage, no consideration is given to pollution by type of industry, or to the abatement of specific pollutants. Rather, the organization of the handbook series has been based on the three basic forms in which pollutants and waste are manifested: gas, solid, and liquid. In addition, noise pollution control is included in the handbook series.

This particular book Volume 9, *Advanced Biological Treatment Processes*, is a sister book to Volume 8 *Biological Treatment Processes.* Both books have been designed to serve as comprehensive biological treatment textbooks as well as wide-ranging reference books. We hope and expect it will prove of equal high value to advanced

undergraduate and graduate students, to designers of water and wastewater treatment systems, and to scientists and researchers. The editors welcome comments from readers in all of these categories.

The editors are pleased to acknowledge the encouragement and support received from their colleagues and the publisher during the conceptual stages of this endeavor. We wish to thank the contributing authors for their time and effort, and for having patiently borne our reviews and numerous queries and comments. We are very grateful to our respective families for their patience and understanding during some rather trying times.

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Principles and Kinetics of Biological Processes

Nazih K. Shammas, Yu Liu, and Lawrence K. Wang

CONTENTS

Abstract Biological technologies can be used to treat a vast majority of organic wastewaters because all organics could be biologically degraded if the proper microbial communities are established, maintained, and controlled. Before environmental engineers design and operate biological treatment systems that create the environment necessary for the effective treatment of wastewater, a sound understanding of the fundamentals of microbial growth and substrate use kinetics is essential. This chapter covers the above including basic microbiology and kinetics, kinetics of activated sludge process, factors affecting the nitrification process, kinetics of the nitrification process, denitrification by suspended growth systems and design examples.

Key Words Activated sludge • biological treatment • denitrification • kinetics • mathematical modeling • allosteric kinetic model • nitrification.

1. INTRODUCTION

Microorganisms are found nearly everywhere in the biosphere and thus are a force in the environment. In the past decades, bacteria have been intensively exploited in wastewater treatment processes. It is therefore the task of the environmental engineer and scientist to understand the role of microorganisms first and then use them to beneficially transform the

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_1

particular environment, such as water or soil. Theoretically, biological technologies can be used to treat a vast majority of organic wastewaters because all organics could be biologically degraded if the proper microbial communities are established, maintained, and controlled. In this regard, many environmental engineering principles have been developed for biological wastewater treatment. Before environmental engineers design and operate biological treatment systems that create the environment necessary for the effective treatment of wastewater, a sound understanding of the fundamentals of microbial growth and substrate utilization kinetics is essential.

2. BASIC MICROBIOLOGY AND KINETICS

Microorganisms are powerful and cheap bioagents of biological wastewater treatment. The performance and stability of a biological treatment system relies on the interaction of different species of living organisms, typically including bacteria, fungi, algae, and protozoa (1).

2.1. Microbial Growth Requirements

Biological processes designed for wastewater treatment must maintain rich microbial populations and enough biomass to metabolize the soluble and colloidal organic wastes. For a successful operation of the biological treatment process, several conditions must be fulfilled, such as the type and concentration of organic waste (as electron donor), electron acceptors, moisture, temperature, necessary nutrients, and the absence of toxic and inhibitory compounds. A sound understanding of these microbial growth requirements is essential for environmental engineers and scientists to design and manage biological wastewater treatment systems.

2.1.1. Electron Acceptors

Aerobic and anaerobic processes are the two main biological technologies used for wastewater treatment. Bacterial respirations for aerobic and anaerobic bacteria need different electron acceptors. The choice of electron acceptors depends on which treatment process is desirable for a specific wastewater (2). For aerobic biodegradation, dissolved oxygen (DO) serves as the terminal electron acceptor. However, under anaerobic conditions, a variety of inorganic compounds can be used as terminal electron acceptors, e.g., NO_3^-, SO_4^{2-}, and so on.

In aerobic systems, the theoretical oxygen demand of an organic compound can be calculated from stoichiometry or determined by laboratory test. The theoretical oxygen demand is the amount of oxygen required to completely oxidize the organic carbon to carbon dioxide and water. As an example, for the complete oxidation of phenol (C_6H_6O) the balanced equation is written as follows:

$$\underset{94}{C_6H_6O} + \underset{224}{7O_2} \rightarrow 6CO_2 + 3H_2O \quad (1)$$

From the molecular weights in Eq. (1) the theoretical oxygen demand of phenol is: $224/94 = 2.38$ mg O_2/mg phenol.

2.1.2. Moisture

Because about 75% of cellular mass is water, and water is a good medium for nutrient transportation, adequate moisture concentration is strongly required in biodegradation of organic chemicals, especially in bioremediation of contaminated soil (3). It is generally accepted that the minimum moisture content necessary for bioremediation of contaminated soil is around 40% of saturation (4). In fact, there is no moisture-associated problem in biological wastewater treatment processes.

2.1.3. Temperature

The performance and response of a biological system depends on temperature variation. The effect of process temperature on microbial activity or the rate of biodegradation can be roughly described by the following simple equation:

$$r_T = r_{20}\alpha^{(T-20)} \tag{2}$$

where

r_T = biodegradation rate at temperature T
r_{20} = biodegradation rate at 20°C
α = temperature-activity coefficient
T = temperature, °C

For most of biological treatment systems, α values are in the range of 1.0 to 1.14 (5). Different groups of bacteria have various temperature optimums. For example, methanogenic bacteria are slow-growing bacteria with a generation time of 3 days at 35°C and 50 days at 10°C, indicating that methane-producing bacteria are very sensitive to changes in temperature (1).

2.1.4. pH

Most bacteria can optimally function only at a relatively narrow pH range of 6 to 8. In biological treatment system, once the reactor pH falls outside the optimal range, the activity of microbial population would drop significantly, and such a decline of activity in turn causes a serious operation problem and may result in the failure of the system (1). Consequently, it is recommended that on-site operators need to regularly monitor the system pH and pay attention to its changes.

2.1.5. Nutrients

Typical elementary composition of bacterial cells based on dry weight is 50% carbon, 20% oxygen, 15% nitrogen, 8% hydrogen, 3% phosphorus and <1% each of sulfur, potassium, sodium, calcium, iron, and magnesium (6). Microbial metabolism requires these elements as nutrients for synthesis and energy generation. The most commonly accepted empirical forms of activated sludge biomass are expressed as $C_5H_7NO_2$ and $C_{42}H_{100}N_{11}O_{13}P$ (7). The empirical formulae of bacterial cells provide a basis for calculation of the N and P requirements for synthesis of biomass from organic waste.

2.2. *Kinetics of Microbial Growth in an Ideal Medium*

Bacteria can grow at high rates under suitable conditions because of their relatively simple structures and growth requirements. However, a particular environment will favor some species more than others.

2.2.1. Kinetics of Microbial Growth

The growth of bacteria in an ideal medium can be described by the best-known Monod equation:

$$\mu = \mu_{\max} \frac{S}{S + K_s} \tag{3}$$

where

μ = specific growth rate
$\mu_{\max}$ = maximum specific growth rate
S = waste concentration
K_s = half-saturation constant

Thus, the rate of bacterial growth in term of mass per unit volume and time can be written as:

$$\frac{dX}{dt} = \mu_{\max} \frac{S}{S + K_s} X \tag{4}$$

where:

X = biomass concentration in the system

In the environmental engineering field, it is accepted that the conversion coefficient of organic waste to new synthesized cells is constant, thus the ratio of the increase in biomass to the decrease in organic substrate is defined as the growth yield coefficient Y,

$$Y = \frac{dX/dt}{dS/dt} \tag{5}$$

Combination of Eqs. (4) and (5) gives the following expression for the rate of waste degradation:

$$\frac{dS}{dt} = \frac{\mu_{\max}}{Y} \frac{S}{S + K_s} X = q_{\max} \frac{S}{S + K_s} X \tag{6}$$

or

$$q = q_{\max} \frac{S}{S + K_s} \tag{7}$$

where,

$q_{\max}$ = maximum specific substrate utilization rate = $\mu_{\max}/Y$
q = specific substrate utilization rate defined as follows:

$$q = \frac{dS}{dt} / X \tag{8}$$

Equation (7) is one of the most commonly used design equations for biological treatment systems. In addition, it can be deduced from above equations that Y can also be defined as the μ/q ratio.

2.2.2. Microbial Decay and Endogenous Respiration

According to Pirt (8), part of the energy source would be used for maintaining the living functions of microorganisms, which is so-called maintenance metabolism. This includes the energy for turnover of cell materials, active transport, motility, and so on. The importance of maintenance metabolism is that the maintenance-associated substrate consumption is not synthesized to new cellular mass. Thus, the biosolids production should be inversely related to the activity of maintenance metabolism (9, 10). On the other hand, to account for the decrease in biomass production that is usually observed when the specific growth rate decreases, Herbert et al. (11) postulated that the maintenance energy requirement could be satisfied through endogenous metabolism. In this case, part of cellular biomass is oxidized to produce the energy for maintenance functions. It is generally assumed that microbial decay occurs following a first-order pattern as follows:

$$\text{Decay rate} = -K_{\text{d}}X \tag{9}$$

where,

K_{d} = constant decay coefficient

Endogenous respiration has profound effect on the production of excessive biosolids. It has been suggested that the aim of both design and operation is to foster as much of this biological decay as possible. Including the decay in Eq. (6) yields an expression for the net growth of biomass in biological system:

$$\frac{\text{d}X}{\text{d}t} = \mu_{\max}\frac{S}{S + K_{\text{s}}}X - K_{\text{d}}X \tag{10}$$

Equations (3) to (10) provide the basis for detailed kinetic analysis and basic design guidelines of biological treatment systems.

2.3. Kinetics of Biological Growth in an Inhibitory Medium

Some substrates may inhibit their own degradation at increased concentrations. When designing and running a biological system for inhibitory waste treatment, environmental engineers must seriously account for the toxicity and inhibition of waste to bacterial growth. It is obvious that the Monod equation does not include the toxic or inhibitory effect, thus they must be modified for biological treatment of inhibitory waste. Figure 1.1 shows typical growth patterns of bacteria in noninhibitory and inhibitory media. It seems that when the concentration of inhibitory substrate is higher than a critical value, a sharp decline in microbial growth is observed, on the other hand, if the concentration of inhibitory substrate is low enough, the inhibitory effect would not be significant.

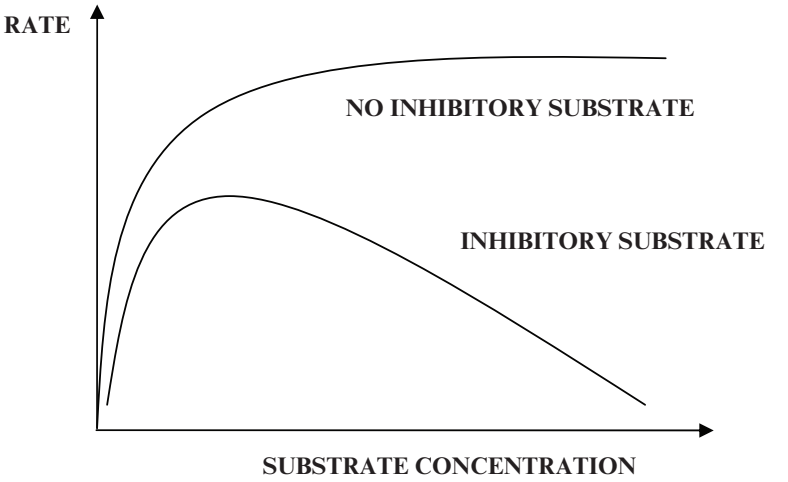

Fig. 1.1. Schematic presentation of inhibitory effect on bacterial growth.

So far, the Haldane equation has been most frequently used to describe the inhibitory effect of a substrate on bacterial growth:

$$\frac{dS}{dt} = q_{max} \frac{S}{S + \frac{S^2}{K_i} + K_s} X \tag{11}$$

where,

K_i = inhibition coefficient

When mixed inhibitory substrates are considered for biological treatment, the expression for inhibitory kinetics will become very complicated because in such a waste mixture, one substrate may inhibit the biodegradation of another.

Current practice shows that for a target inhibitory substrate, its concentration is critical for biological treatment. If the threshold of substrate concentration that bacteria can bear is exceeded, inhibition, and die-off of bacteria in the reactor will start on a continuing and irreversible basis, leading to serious loss or even failure of the system's purification efficiency and capability. Predetermination of inhibitory threshold of substrate concentration is essential for the design of a biological treatment system for inhibitory wastes. In industrial practice, where inhibitory wastes are more common, there are some technical measures that can help to mitigate inhibition, such as acclimation of bacteria, introduction of robust species, or dilution of the waste stream.

2.4. Minimum Substrate Concentration

In many cases, the characteristics of soluble wastes found in soil and wastewater have dual effects on biological treatment processes; one, when the concentrations of waste constituents are generally low and two, when their toxicity to microbial activity is relatively high. A low

waste concentration may be risky in case it could not support a sustainable and viable biomass needed for biological treatment. As Eq. (3) indicates, the specific growth rate of microorganisms is proportionally related to substrate concentration. Microbial growth could cease as the substrate concentration diminishes to a certain low unsustainable concentration. For a biological treatment system, a minimum substrate concentration is required to sustain a viable biomass. In the environmental engineering field, the minimum substrate concentration ($S_{\min}$) is defined as the substrate concentration at which formation of new biomass equals its loss by endogenous respiration (3). When the minimum substrate concentration occurs, Eq. (10) shows that

$$\mu_{\max}\frac{S_{\min}}{S_{\min}+K_s}X - K_dX = 0 \tag{12}$$

that is,

$$S_{\min} = \frac{K_sK_d}{\mu_{\max}-K_d} \tag{13}$$

or

$$S_{\min} = \frac{K_sK_d}{Yq_{\max}-K_d} \tag{14}$$

2.5. *Mathematical Approximation for Wastewater Treatment*

In many situations of wastewater treatment, a simple first-order approximation has been used with reasonable accuracy to describe the biodegradation of organic wastewater. This approximation is based on two main assumptions (4):

1. The target substrate or waste is at a relatively low concentration.
2. The biomass concentration in the system is at a steady state, consequently it changes little with operation time and can be regarded as a constant.

Thus, Eq. (6) reduces to:

$$\frac{dS}{dt} = \frac{q_{\max}}{K_s}XS \tag{15}$$

or

$$\frac{dS}{dt} = k_1S \tag{16}$$

where,

$k_1 = Xq_{\max}/K_s$ = first-order biodegradation rate constant

Integrating both sides of Eq. (16) yields

$$S = S_oe^{-k_1t} \tag{17}$$

where,

t = reaction time

S_o = initial substrate concentration at $t = 0$

S = substrate concentration at any time t

In case the substrate concentration is relatively higher than K_s and X is considered constant, Eq. (6) can be simplified to

$$\frac{dS}{dt} = q_{max} X \tag{18}$$

Equation (18) shows a zero-order reaction, that is,

$$S = S_o - k_o t \tag{19}$$

where,

$k_o = X q_{max}$ = zero-order rate constant

Reported examples of zero-order biodegradation kinetics include substances such as glucose, phenol, phthalic acid, aspartic acid, ethanol, and acetate (5).

3. KINETICS OF ACTIVATED SLUDGE PROCESSES

3.1. Brief Description of Activated Sludge Processes

The activated sludge process is the most widely used biological process for treatment of a variety of wastewaters. In the past century many modifications of the basic activated sludge process have evolved for various purposes (2):

1. Complete-mix activated sludge process: A completely mixed system can allow a more uniform aeration of the wastewater in the aeration tank. This process has been applied to handle a variety of wastewaters with great success, especially because the process can sustain shock and toxic loads.
2. Step-aeration activated sludge process: In this modified system, influent wastewater is distributed through several points in the aeration tank. This leads to a relatively homogenous load distribution along the length of the aeration tank resulting in a more efficient use of dissolved oxygen.
3. Contact-stabilization activated sludge: The influent contacts with a high concentration of biomass in a small contact tank for a short period of time (20 to 40 min). The mixture then flows to the secondary clarifier where it gets settled and the resulting biosolids are returned to a stabilization tank with a hydraulic retention time of 4 to 8 h. In this contact tank, a rapid biosorption of organic compounds is expected followed by the oxidation of the organics. This system would need smaller tankage and produce smaller amounts of biosolids.
4. Tapered aeration process: In the basic activated sludge process, organic influent is one-point loaded to the head of aeration tank, thus the oxygen demand is extremely high at the head of the aeration tank, but very low at the exit end. To overcome this problem, in tapered aeration process, the air supply tapers off with distance along the aeration tank so that supply and demand can be balanced throughout the tank.
5. Pure oxygen activated sludge process: The pure oxygen activated sludge process is based on such a simple idea that the rate of oxygen transfer in water is proportional to the partial pressure of oxygen, that is, the rate of oxygen transfer is higher for pure oxygen than for atmospheric

oxygen. Higher availability of oxygen for microorganisms leads to improved treatment efficiency and reduced production of biosolids and reactor volume.

3.2. *Kinetics of Completely Mixed Activated Sludge Process*

3.2.1. Basic Design Models

Development of design models for a completely mixed activated sludge process is based on a mass balances on microorganisms and substrate around the system together with the kinetics of microbial growth and waste utilization Figure 1.2 shows the items of interest in the mass balance.

To develop a mass balance for the reactor system, some basic assumptions are made (12):

1. Biodegradation of organic wastes takes place only in the aeration tank.
2. No biological reactions take place in the settling tank and the biomass in the secondary clarifier is negligible.
3. No active biomass is present in the influent to the aeration tank.
4. The substrate is soluble so that it cannot settle out in the secondary clarifier.

In Fig. 1.2, X, X_r, X_w, and X_e represent the active biomass concentrations. The following is based on the work by Lawrence and McCarty (12). The mass balance on biosolids is expressed as follows:

$$\text{Accumulation} = \text{in} - \text{out} + \text{generation} \tag{20}$$

According to Eq. (20), the mass balance for microorganisms around the whole system is:

$$V\frac{dX}{dt} = 0 - (Q_e X_e + Q_w X_w) + [Y(r_s)V - K_d X V] \tag{21}$$

where,

r_s = rate of soluble substrate utilization
V = Volume
Q = flow rate

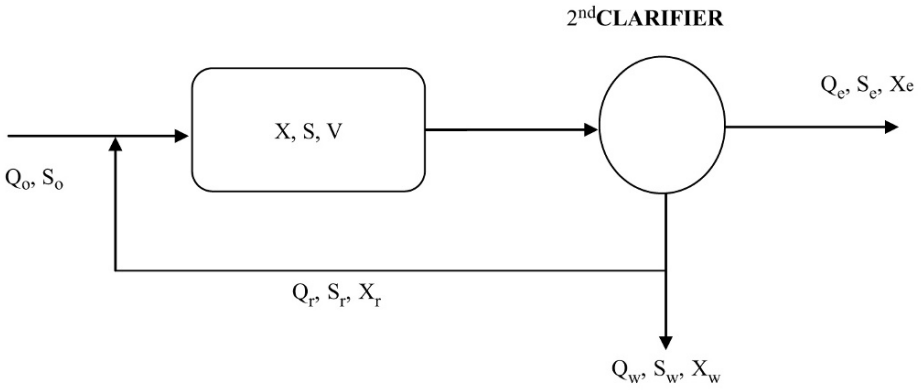

Fig. 1.2. Mass flow chart of completely mixed activated sludge process.

Similarly, the mass balance on substrate yields:

$$V\frac{\mathrm{d}S}{\mathrm{d}t} = Q_o S_o - (Q_\mathrm{e} S_\mathrm{e} + Q_\mathrm{w} S_\mathrm{w}) - r_\mathrm{s} V \tag{22}$$

It should be pointed out that Eqs. (21) and (22) are derived on the basis of a mass balance on biomass and substrate, respectively, thus can be used to describe the operation of the system under nonsteady or steady-state conditions. In practice, activated sludge processes are run under steady-state conditions. At steady state, the changes in accumulation of both biomass and substrate are zero, that is,

$$V\frac{\mathrm{d}X}{\mathrm{d}t} = 0 \text{ and } V\frac{\mathrm{d}S}{\mathrm{d}t} = 0 \tag{23}$$

To facilitate the development of a design model, we need to define some very useful operation and control parameters:

Mean hydraulic retention time for the aeration tank (θ):

$$\theta = \frac{V}{Q_o} \tag{24}$$

Mean cell retention time or solids retention time (θ_x):

$$\theta_x = \frac{\text{biomass in the aeration tank}}{\text{discharge rate of biomass}} \tag{25}$$

That is,

$$\theta_x = \frac{VX}{X_\mathrm{e} Q_\mathrm{e} + X_\mathrm{w} Q_\mathrm{w}} \tag{26}$$

At steady state, Eq. (21) can be rearranged as follows:

$$\frac{X_\mathrm{e} Q_\mathrm{e} + X_\mathrm{w} Q_\mathrm{w}}{XV} = \frac{Y(r_\mathrm{s})}{X} - K_\mathrm{d} \tag{27}$$

Comparing Eqs. (26) and (27) one can deduce that:

$$\frac{1}{\theta_x} = \frac{Y(r_\mathrm{s})}{X} - K_\mathrm{d} \tag{28}$$

Equation (28) is an important design relationship for the completely mixed activated sludge process. It can be applied whatever the form of r_s may be; a Monod equation, a first-order approximation for dilute wastewater or the Haldane equation for high-concentration inhibitory organics. If we assume that for a wastewater the Monod equation is applicable, then

$$\frac{1}{\theta_x} = Y\frac{q_\mathrm{max} S_\mathrm{e}}{S_\mathrm{e} + K_\mathrm{s}} - K_\mathrm{d} \tag{29}$$

Solving Eq. (29) for S_e gives:

$$S_\mathrm{e} = K_\mathrm{s}\frac{1 + K_\mathrm{d}\theta_x}{\theta_x(Y q_\mathrm{max} - K_\mathrm{d}) - 1} \tag{30}$$

Equation (30) is one of the recognized design equations originally derived by Lawrence and McCarty (12). This equation shows that the efficiency of substrate removal is proportional to the sludge age. Thus, environmental engineers can expect to need a relatively large θ_x to obtain high treatment efficiency; while at the same time have a short hydraulic retention time, which means a small reactor volume.

Similarly, at steady state, Eq. (22) can be rearranged to give r_s as a function of S:

$$r_s = \frac{Q_o S_o - Q_e S_e - Q_w S_w}{V} \tag{31}$$

The substrate concentration in the aeration tank, S, is equal to the concentration in the effluent S_e as well as in the waste sludge line, S_w because no biological reaction occurs in the settling tank. Also from the continuity equation of fluid flows one can state that:

$$Q_o = Q_e + Q_w \tag{32}$$

Using the above relationships, Eq. (31) becomes

$$r_s = \frac{Q_o(S_o - S_e)}{V} = \frac{S_o - S_e}{\theta} \tag{33}$$

Like Eq. (30), Eq. (33) is another general representation of the waste removal rate in terms of system characteristics. Substitution of Eq. (33) into (27) produces,

$$X = \frac{\theta_x}{\theta} \frac{Y(S_o - S_e)}{1 + K_d \theta_x} \tag{34}$$

Equation (34) indicates that the biomass concentration in the aeration tank depends on the ratio of solids retention time to the hydraulic retention time, θ_x/θ. This equation is one of the most commonly recognized design formulas (7, 12).

3.2.2. Process Control Parameters

Equations (30) and (34) can be useful in predicting the effects of various changes in system parameters, but they are difficult to use from a design standpoint because of the many kinetic constants involved. Environmental engineers and scientists have developed more usable process design relationships enthatough are widely used in process design practice. These include the specific removal rate of soluble waste (q), mean solid retention time (θ_x), and the food-to-microorganisms, F/M, ratio (7). The following discussion is based on material from Metcalf and Eddy (7).

The specific removal rate of soluble waste, q: The specific removal rate of soluble wastes is defined as,

$$q = \frac{r_s}{X} = \frac{S_o - S_e}{\theta X} = \frac{Q_o}{V} \frac{S_o - S_e}{X} \tag{35}$$

To determine q, the liquid waste flow and the biomass effective in substrate utilized must be known. The substrate utilized can be quantified by the difference between the influent and effluent waste concentrations ($S_o - S_e$). However, the evaluation of the active biomass

of microorganisms, X, is not an easy task, which in practice can be roughly quantified by measuring the mixed liquor volatile suspended solids (MLVSS) in the aeration tank.

Solid retention time or sludge age (θ_x): θ_x is defined by the expression in Eq. (25). In the completely mixed activated sludge process with sludge recycle, as shown in Fig. 1.2, excessive sludge wastage can be accomplished by directly discharging from the aeration tank or wasting from the mixed-liquor return line. In practice, to obtain a thicker sludge, wasting is preferred by drawing off sludge from the recycle line (2). If the system is operated correctly, X_r (which is equal to X_w) is much larger than X_e, thus Eq. (26) can be simplified to

$$\theta_x \approx \frac{VX}{X_r Q_w} \tag{36}$$

Equation (36) shows that to control sludge age the biomass concentrations in both aeration tank and return sludge line must be known. The biomass concentration in the return line (X_r) can be roughly estimated in the following way:

$$X_r = \frac{10^6}{\text{SVI}} \tag{37}$$

where

X_r = biomass concentration in the return line (mg/L)
SVI = sludge volume index

The sludge volume index (SVI) is a measure of the ability of sludge to settle and compact, which can be easily determined from a laboratory column settling test (13). SVI is defined as the volume in ML occupied by 1 g of activated sludge mixed liquor solids, dry weight, after settling for 30 min in a 1-L graduated cylinder (16).

θ_x indeed describes the residence time of the sludge in the aeration tank. The sludge requires a certain time to assimilate the liquid waste and reproduce itself. If the sludge is not able to reproduce itself before being washed out of the aeration tank, the operation will fail. On the other hand, higher sludge age may cause the sludge to undergo more endogenous decay leading to poorer settleability of the sludge and effluent quality. The control of θ_x means the control of the sludge growth rate, and hence the degree of waste stabilization (2). To maintain a desirable sludge age, a specific percentage of the biomass in the system must be wasted daily. Substituting Eq. (35) into (28) gives

$$\frac{1}{\theta_x} = Yq - K_d \tag{38}$$

Equation (38) reveals a direct relationship between the net specific growth rate, $1/\theta_x$, and the specific removal rate of liquid waste, q. In addition, when the effect of endogenous respiration on the true growth yield (Y) is taken into account, the observed growth yield (Y_{obs}) of biomass is lower than Y and can be expressed as

$$Y_{obs} = \frac{Y}{1 + K_d \theta_x} \tag{39}$$

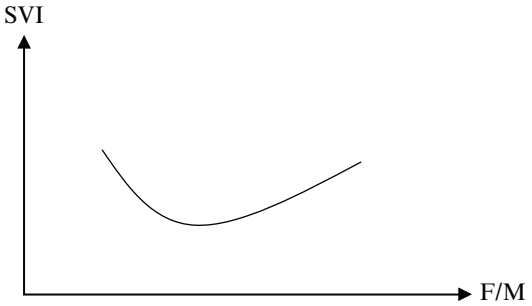

Fig. 1.3. Effect of F/M ratio on SVI of biosolids. (*Source*: Adapted from (6)).

Food to microorganisms' ratio (F/M ratio): In the environmental engineering field, food to microorganisms' ratio (F/M) is defined as

$$\mathrm{F/M} = \frac{S_o}{\theta X} \tag{40}$$

The physical meaning of this parameter indeed describes the degree of starvation of the sludge or the potential food availability to the sludge in the system. It is known that the F/M ratio influences the ability of the sludge to swttle and compact. A typical plot of SVI against F/M ratio is presented in Fig. 1.3.

3.2.3. Process Management

For a completely mixed activated sludge process, the performance and stability of the system is highly dependent on the system sludge age. For a target waste, a given biological community and the known environmental conditions, the kinetic constants in Eq. (30), Y, $q_{\max}$, K_s, and K_d are fixed. In this case, Eq. (26) clearly shows that the target waste concentration in effluent (S_e) is a function of the sludge age (θ_x). A schematic presentation of Eq. (26) plotted as S_e versus θ_x is shown in Fig. 1.4. The figure reveals that there exists a critical value of the sludge age below which waste biodegradation does not occur. This critical value of θ_x is then defined as the minimum sludge age or minimum solid retention time $(\theta_x)_{\min}$. The physical meaning of this parameter is that $(\theta_x)_{\min}$ reflects the retention time of sludge at which the biomass is washed out or wasted from the system faster than it can be reproduced (7). It seems from Fig. 1.4 that when washout occurs, the influent waste concentration (S_o) should equal the waste concentration in effluent (S), hence the minimum sludge retention time (SRT) or sludge age can be calculated using Eq. (29), that is,

$$\frac{1}{(\theta_x)_{\min}} = Y\frac{q_{\max}S_o}{S_o + K_s} - K_d \tag{41}$$

In many actual operation cases, S_o is usually much greater than K_s. Hence, Eq. (41) can be reduced to

$$\frac{1}{(\theta_x)_{\min}} = Yq_{\max} - K_d \tag{42}$$

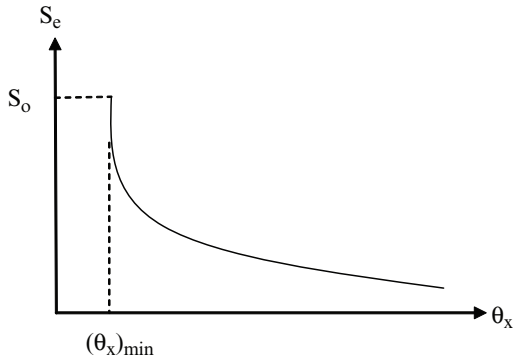

Fig. 1.4. Relationship between effluent concentration and biosolids age. (*Source*: Adapted from (14)).

It must be stressed that a biological treatment system requiring a certain target effluent concentration must be designed with θ_x greater than its minimum value. According to Eckenfelder and Argaman (14), in real system design, a safety factor of 2 to 20 is usually considered. Hence,

$$\theta_x = SF \times (\theta_x)_{\min} \tag{43}$$

where,

SF = safety factor

$(\theta_x)_{\min}$ = minimum value of sludge age or sludge retention time (SRT)

For a given wastewater, many factors may affect the selection of SF. Such factors include fluctuations in operation temperature, in wastewater flow rate and in wastewater strength and characteristics: desired treatment efficiency; required reliability in operation; reactor configuration and nutrient removal.

In addition, microorganisms are the main agents for the bio-oxidation of organics, thus biomass concentration in the aeration tank is another key factor for maintaining the stability of the system. The maintenance of suspended solids is dependent, to a great extent, upon the settleability and recycling extent of the sludge. The recycle ratio of sludge from the secondary clarifier is defined as

$$R = \frac{Q_{\mathrm{r}}}{Q_o} \tag{44}$$

where,

R = recycling ratio

A biomass balance around the aeration tank gives

$$X = X_{\mathrm{r}} \frac{R}{1+R} \tag{45}$$

or

$$R = \frac{X}{X_r - X} \tag{46}$$

3.3. Oxygen Requirements

Air is supplied to the aeration tank to satisfy the biochemical oxygen demand (BOD) in the process of organic oxidation. In addition, diffused air is used for turbulent mixing to keep the biological sludge in suspension and provide initial contact with the substrate. This is particularly true for diffused aeration although mechanical aeration provides good mixing without relying on the diffused air in the wastewater. It is believed also that turbulent mixing by diffused air facilitates mass transfer of oxygen into the biological flocs and transfer of carbon dioxide and other waste products out of the flocs. In the activated sludge process, the oxygen requirement consists of the amount of oxygen needed for both synthesis and respiration. Consequently one needs to know the ultimate BOD of the wastewater that can be calculated from BOD_5 using an appropriate conversion factor. The respiration oxygen demand is 1.42 g O_2/g MLVSS (15). Because part of the MLVSS produced is wasted in the process operation for the control of sludge retention time, the respiration oxygen demand is reduced by an amount proportional to the amount of wasted sludge. According to Wang (16), the theoretical oxygen requirement for an activated sludge process therefore is:

$$\text{Daily theoretical } O_2 \text{ requirement} = \text{BOD removed daily} - 1.42\ (\text{VSS wasted daily}) \tag{47}$$

in which all terms are expressed in mass per day. In practice, air is supplied to the aeration tank mixed liquid to maintain a minimum dissolved oxygen concentration of 1 to 2 mg/L. The objective is to maintain a dissolved oxygen gradient across the liquid–floc interface to ensure an effective oxygen transfer into the biological flocs. The critical oxygen tension for the biological floc is believed to be in the neighborhood of 0.1 mg DO/L. Equation (47) can be used for the calculation of theoretical oxygen requirements of an activated sludge system. In practice, oxygen uptake rate (OUR) is a useful process control parameter. Any changes in OUR reflects the need for a change in operation (6, 17).

3.4. Biosolids Production

The activated sludge process has been applied worldwide in municipal and industrial wastewater treatment practice. Removal of organic materials by biological oxidation is a core technology in wastewater treatment processes. New biomass, carbon dioxide, soluble microbial products, and water are the end products for this process. The daily production of excess biosolids from a conventional activated sludge process is around 15 to 100 L/kg of BOD_5 removed, out of which more than 98% is water (18). For an activated sludge process control, it is important to know the quantity of excess biosolids to be produced daily, as it will affect the design of the biosolids treatment facilities. As discussed earlier, the rate of change of biomass concentration in a reactor, $V(dX/dt)$, is equal to the net rate of microbial growth in the reactor, $V(YXq_{max} - K_dX)$, minus the rate of biomass outflow from the system. Therefore, to maintain a constant biomass concentration in the aeration tank, the excess

biosolids production rate on a mass basis must be equal to $V(YXq_{max} - K_dX)$. The following equation describes the above situation (19):

$$V(dX/dt)_{excess} = Q_wX_w = V(Yq_{max}X - K_dX) = VX/\theta_x \quad (48)$$

where,

$V(dX/dt)_{excess}$ = excess biosolids production rate
X_w = wasted biosolids concentration
Q_w = wasted biosolids flow rate

Many operating parameters can affect the production of excess biosolids from a biological treatment process. These include sludge age, temperature, and dissolved oxygen concentration.

Different opinions can be found in the literature with regard to the effect of dissolved oxygen concentration on biosolids production (7, 20–22). It is generally recognized that in an activated sludge process, supply of dissolved oxygen plays a limiting role on any future increase in the loading rate on the treatment facility. Results from purified oxygenation activated sludge process show that the growth yield can be lowered by up to 54% as compared with conventional air-activated sludge system even at high biosolids loading rate (20). Boon and Burgess (23) compared the biosolids production in oxygen and air-activated sludge systems. They found that for similar biosolids retention time, the observed biosolids yield in the pure oxygen system was only 60% of that in the air system. Abbassi et al. (22) also reported that the excess biosolids production decreased from 0.28 mg MLSS/mg BOD_5 to 0.20 mg MLSS/mg BOD_5 as the reactor DO was increased from 1.8 to 6.0 mg/L in a laboratory-scale conventional activated sludge reactor.

In the current activated sludge theory, sludge age (θ_x) is defined as the average time a unit of biomass remains in the treatment system. Much research has shown that θ_x is the most important operational parameter in the activated sludge process. For a steady state system, the θ_x is inversely related to the specific growth rate. It has been demonstrated that the relationship between the observed sludge yield (Y_{obs}) and sludge age can be described by the following expression (12):

$$\frac{1}{Y_{obs}} = \frac{1}{Y_{max}} + \frac{\theta_x K_d}{Y_{max}} \quad (49)$$

where,

Y_{max} is the maximum growth yield

Equation (49) shows that the observed growth yield is inversely proportional to sludge retention time and endogenous decay rate in a steady state activated sludge process. This equation also provides a theoretical basis for in-plant engineers to control the total biosolids production by adjusting θ_c during the wastewater biological treatment. Stall and Sherrard (24) reported that excess biosolids production was reduced by 60% when the θ_x was increased from 2 to 18 days, while no effect on COD removal efficiency was observed. On the other hand,

Wunderlich et al. (25) showed that in a high-purify oxygen activated sludge system, biosolids production was reduced from 0.38 to 0.28 mg MLVSS/mg COD removed as the θ_x increased from 3.7 to 8.7 days. It seems from these results that the pure oxygen aeration process operated at a relatively long θ_x would be much more beneficial to the reduction of excessive biosolids production.

The general purpose of the activated sludge process is the removal of organic pollutants rather than the cultivation of excess biosolids. With increase of population and expansion in industrialization, the management of the increased excess in biosolids production is generating a real challenge in the field of environmental engineering. So far the regulations in biosolids management in most countries are becoming more and more stringent in relation to the application of biosolids on agricultural land, dumping into sea, or disposal in landfill. Waste activated sludge production is an important economic factor because the generated biosolids have to be treated before reuse or disposal in an environmentally sound and cost-effective manner. The treatment of excess biosolids may account for 25% up to 65% of a total plant operation cost (26, 27). Also it is necessary to look for appropriate ways to recycle the excess biosolids production for beneficial uses. Hence, an ideal way to solve the biosolids-associated problems is to reduce their production in first place rather than spending valuable resources in post-treatment of the generated product (27). Strategy for minimization of excess biosolids production from biological treatment processes has become a very practical and urgent issue (28).

4. FACTORS AFFECTING THE NITRIFICATION PROCESS

The Michigan studies on the significance of nitrogenous oxidation in creating oxygen sag in receiving streams and other studies showing the role of ammonia and nitrate nitrogen in stimulating algal blooms have demonstrated the need for information on how wastewater-treatment plants can be designed to optimize nitrification and denitrification processes.

Nitrogen removal from wastewater can be accomplished through a variety of alternative processes. The popular approach is by biological nitrification-denitrification (29–37), which has the additional advantage of returning nitrogen to the atmosphere in its natural form. In this regard, it has been shown that the efficiency of nitrogen removal is strictly correlated with the degree of nitrification achieved (31). Moreover, the process of denitrification is quite effective and the nitrification phase is the limiting step in determining the efficiency of nitrogen extraction. It can be concluded that further perfection of the overall process depends on the improvement of the nitrification phase, which is the less reliable phase in the process sequence. In simple terms, nitrification in treatment plants can be maintained only when the rate of growth of nitrifying bacteria is rapid enough to replace organisms lost through biosolids wasting. When these bacteria can no longer keep pace, the ability to nitrify decreases and may become extinct.

To be able to evaluate accurately the effect of the environmental factors and to present a consistent and valuable basis for application, it is clear that a kinetic description of the process is essential. Several equations have been proposed to describe the nitrification process (38).

The kinetic expression most extensively used to describe biological systems is the one postulated and experimentally sustained by Monod (39–43). Eqs. (6) and (7) discussed in a previous section can be expressed in the following form:

$$v = \frac{dS}{dt} = kX\frac{S}{S + K_s} = V_m\frac{S}{S + K_s} \tag{50}$$

where

$v = (qX)$ = rate of substrate (NH_3-N) utilization, mg/L/d
S = substrate ammonia nitrogen concentration, mg/L
t = time, day
dS/dt = rate of substrate (NH_3-N) utilization, mg/L/day
$k = (q_{max})$ = rate of NH_3-N utilization per unit weight of microorganisms, mg/L NH3-N/mg/L MLVSS/day
X = concentration of microorganisms, MLVSS, mg/L
K_s = half-velocity coefficient, mg/L
$V_m = kX = (q_{max}\ X)$ maximum rate of ammonia utilization, mg/L NH_3-N/day or mg/L/day

The inverse of the above equation is shown in Eq. (51), which plots as a straight line when $1/v$ is drawn versus $1/S$.

$$\frac{1}{v} = \frac{1}{dS/dt} = \frac{1}{kX} + \frac{K_s}{kX}\frac{1}{S} = \frac{1}{V_m} + \frac{K_s}{V_m}\frac{1}{S} \tag{51}$$

Shammas (29, 44) carried out an extensive and systematic research involving 45 separate experimental studies under various controlled operational conditions to determine the best suitable kinetic model for the nitrification process (more on kinetic modeling in Section 5). The first step for evaluating the kinetic parameters V_m, K_s, and k is to determine the nitrification rate, v, as a function of substrate concentration. From plots of ammonia-nitrogen concentration versus time (0 to 8 h) for all 45 experiments, values of v were determined from the slopes of the tangents at different substrate concentrations. K_s and V_m (hence k) were determined from the reciprocal plots of $1/v$ against $1/S$, taking advantage of the linearity of the plots at high values of v. The intercepts on the $1/v$ axis yield the values of V_m (hence k); the values of K_s are obtained from the slopes (44).

4.1. Factors Affecting the Half-Velocity Coefficient, $\mathbf{K_s}$

The variation of this parameter with temperature and pH at different MLVSS concentrations is shown in Table 1.1. At low (430 mg/L) MLVSS, K_s decreases with increasing temperature (4°C to 33°C) and pH (7.0 to 8.3). While the K_s values for higher MLVSS concentrations also tend to decrease with increasing pH and temperatures to 10°C and 17°C, the trend reverses itself at higher temperatures. This reversal seems to begin at 10°C to 20°C, with a small variation in K_s after 25°C (44).

An interesting feature of this change in behavior created by the increase in microbial mass is that it altered both the pH and temperature effects on K_s (44). The shift is much more

Table 1.1
Variation of K_s with pH, temperature, and MLVSS (44)

		K_s (mg/L)		
		Nominal MLVSS (mg/L)		
pH	T (°C)	430	1200	3200
7.0	4	22	20	9.5
	10	19	15	3.8
	17	16	1.5	7.4
	25	5.7	3.4	8.0
	33	4.7	4.0	8.0
7.7	4	20	19	14
	10	10	8.8	3.0
	17	8.8	4.3	2.2
	25	5.6	5.7	12
	33	3.8	6.8	12
8.3	4	14	13	7.3
	10	7.2	7.0	2.5
	17	5.4	4.8	3.0
	25	4.4	6.2	16
	33	3.5	7.0	16

pronounced at the highest MLVSS concentration. For a microbial mass of 3,200 mg/L the values of K_s at 25°C and 33°C were far higher than those at 10°C and 17°C. The minimum values of K_s (which correspond to higher oxidation velocities) at 430, 1,200, and 3,200 mg/L MLVSS occur at temperatures of 25°C to 33°C, 17°C, and 10°C to 17°C, respectively, for all pH values. Low K_s values for 430 mg/L MLVSS occur at pH 8.3 for all temperatures, whereas at the higher concentrations (1,200 and 3,200 mg/L MLVSS), low values of K_s occur at either pH 8.3 or 7.0, depending on whether the temperature is below or above 17°C.

Thus, there is a significant interaction between pH and temperature and MLVSS. Each of the parameters studied affects the K_s value in a way that depends on the biomass concentration (44). This behavior may explain the variations in values of design parameters reported by authors using several different rate equations (38, 40). Downing et al. (45) found that the half-velocity coefficient was usually very small, thus rendering the Monod model close to a zero-order reaction (not substrate limiting). This was supported by Knowles et al. (46) who stated that K_s is 0.2 to 1.7 mg/L for Nitrosomonas and 0.18 to 0.25 mg/L for Nitrobacter. On the other hand, Stratton and McCarty (47) showed that K_s ranged from 1.25 to 5.59 mg/L for ammonia-nitrogen oxidation. Similarly, Painter (13) reported that the K_s values for ammonia-nitrogen oxidation varied between 1.0 mg/L at 20°C to 3.5 mg/L at 25° and 10 mg,/L at 30°C. For nitrite-nitrogen oxidation, the coefficient changed from 5 mg/L at 25°C to 8.4 mg/L at 32°C. Randall and Buth (40) demonstrated that the nitrification rate changes from zero-order

to a higher order simply because of temperature changes. In this study (44), the sensitive temperature range at which this occurs is between 10°C and 17°C. This may very well explain the different reaction orders and constants reported by the other researchers.

4.2. *Factors Affecting the Maximum Rate Constant, k.*

The values of the rate constant were 0.0085/day at 4°C and pH 7, to 0.175/day at 33°C and pH 8.3. These values correlate well with the results of different studies on nitrifier-enriched activated sludge. However, they are much lower than those reported for either river water or pure culture. This should be expected because the rate in activated sludge is based on MLVSS, which, even when enriched with nitrifiers, is not completely composed of nitrifying organisms. Painter and Jones (48) reported that the highest rate they could obtain was 0.144/day and that the rate was usually between 0.05 and 0.07/day; the maximum rate constant was only about 2% of that of a pure culture. Wild et al. (49) found that the rate varied from 0.185/d at pH 8.4 to a minimum of 0.020/d at pH 6. Bishop et al. (50) reported a rate of 0.11/d at 27°C, which decreased to 0.032/d at 15°C. Sutton et al. (39) demonstrated that at a MLVSS concentration of 1,700 mg/L, pH 7 to 8, and 21°C, k was 0.0216/d. They also reported that the sludge retention time had to be doubled from 30 to 60 ds to attain the same extent of nitrification at 10°C.

The results of the effect of temperature and pH on k at different MLVSS concentrations are shown in Figs. 1.5 to 1.7. The data from each of the 45 runs are depicted by three sets of curves. Each set represents the variation of k with temperature at three different pH values and for a particular MLVSS concentration (44).

The linear fit of the data indicates that the maximum rate constant varies logarithmically with temperature from 4° to 33°C in the case of low solids concentration, and from 4° to 25°C for high MLVSS concentrations. The parallel regression lines for the different pH values for each of the MLVSS concentrations indicate that pH and temperature do not interact but affect the maximum nitrification rate constant independently (44). However, because the slope of each set of lines is different at each of the MLVSS values, it follows that MLVSS concentration has an influence on the extent of temperature and pH effects.

The equation of the regression line fitted to the data is similar to the popular modified Arrhenius relationship, Eq. (2):

$$k = k_{20}\, e^{b(T-20)} \tag{52}$$

where

k = maximum rate constant at temperature T (1/d)
k_{20} = maximum rate constant at 20°C (1/d)
T = temperature (°C)
b = temperature coefficient ($e^b = \alpha$)

From the above discussion, and as indicated in Figs. 1.5 to 1.7, the value of b is constant with respect to pH, and variable with MLVSS concentration. Figure 1.8 shows a log–log plot of the temperature coefficient (b) against the MLVSS concentration (X). The equation of the fitted line is (44):

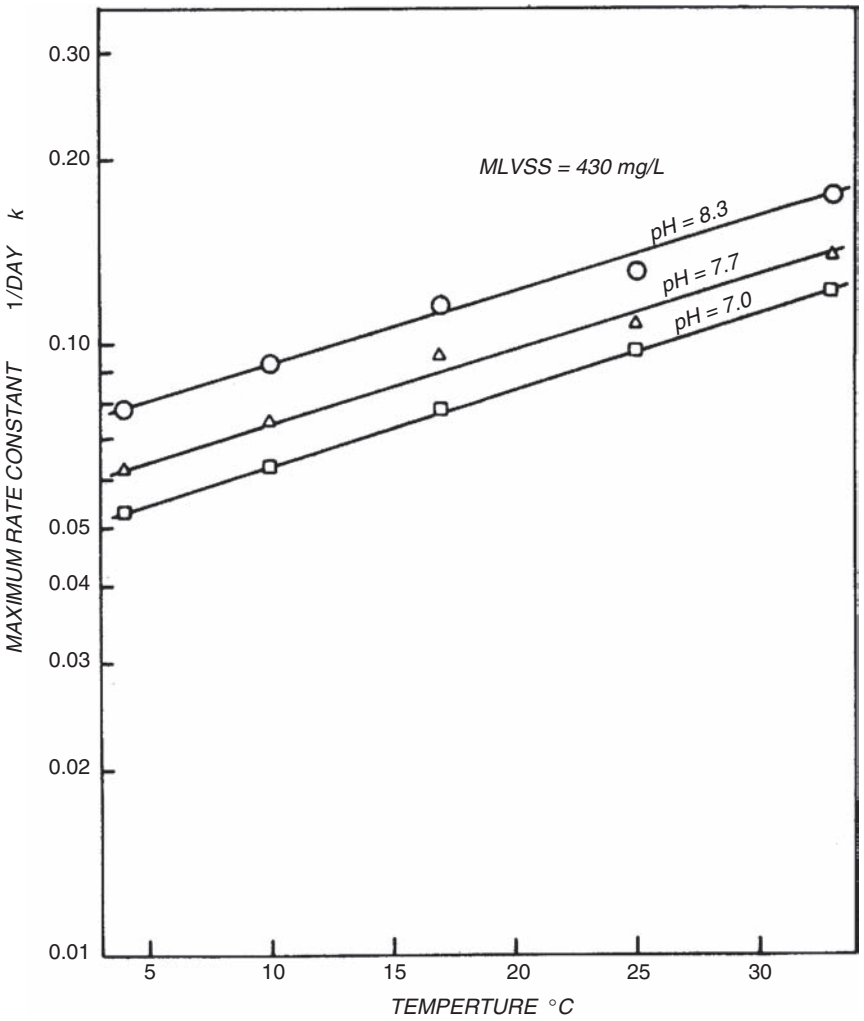

Fig. 1.5. Variation of rate constant with temperature at different pH values (44).

$$b = 0.00044 \times^{0.69} \tag{53}$$

or, in terms of a known value b_1 at X_1,

$$b = b_1 \left(\frac{X}{X_1} \right)^{0.69} \tag{54}$$

The values of b are presented in Table 1.2 with comparable values from other sources. This table shows that temperature coefficients reported in different studies under different conditions fall within the range of coefficients determined in this study (44) The highest coefficient

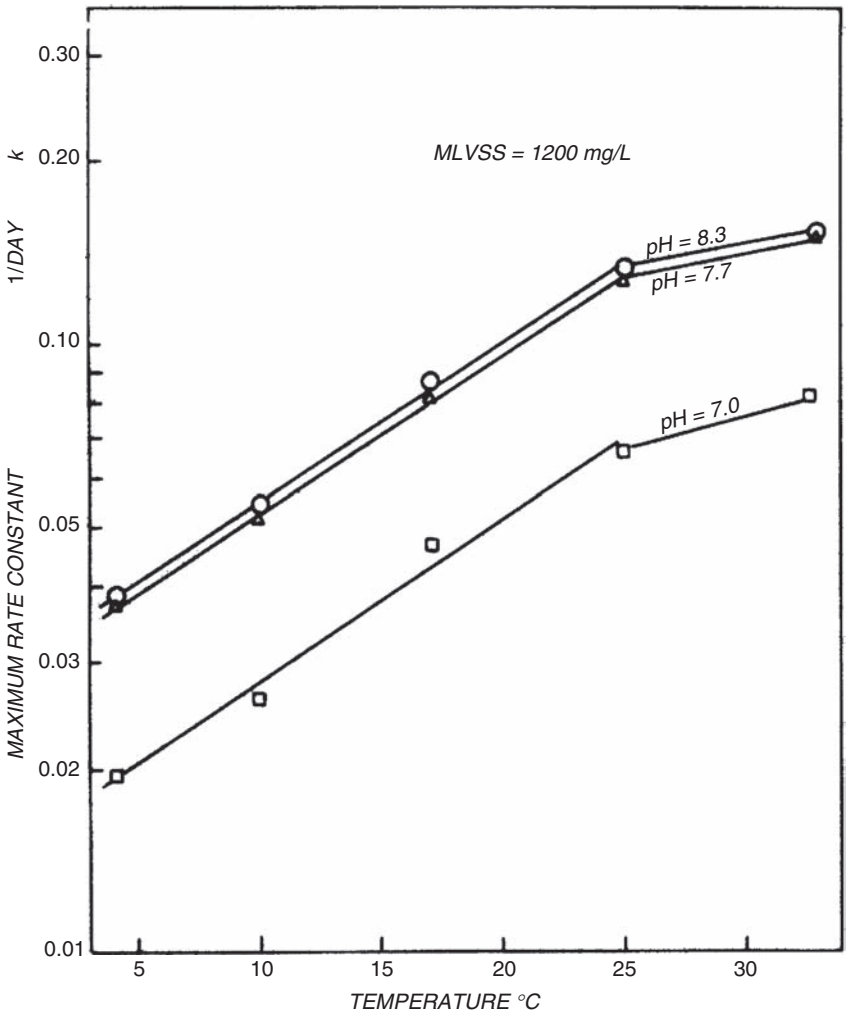

Fig. 1.6. Variation of rate constant with temperature at different pH values (44).

for ammonia oxidation in an activated sludge medium (0.12) was reported by Downing (51). This value corresponds to the coefficient determined in this study for 3,200 mg/L MLVSS. The lowest temperature coefficients for *Nitrobacter* (0.056 and 0.059) reported by Stratton and McCarty (47) and Knowles et al. (46) are equivalent to the value for 1200 mg/L MLVSS. Other reported coefficients, 0.073 (52), 0.075 (33, 35), 0.084 (47), and 0.095 (46), were scattered among these maximum and minimum values.

Although the maximum rate constant k is supposedly independent of the MLVSS concentration, the variation in the temperature coefficient with MLVSS, indicated by the slopes of the lines in Figs. 1.5 to 1.7, implies that this is not always the case. To clarify this behavior, the maximum nitrification velocity (V_m) was plotted against temperature in Fig. 1.9. The

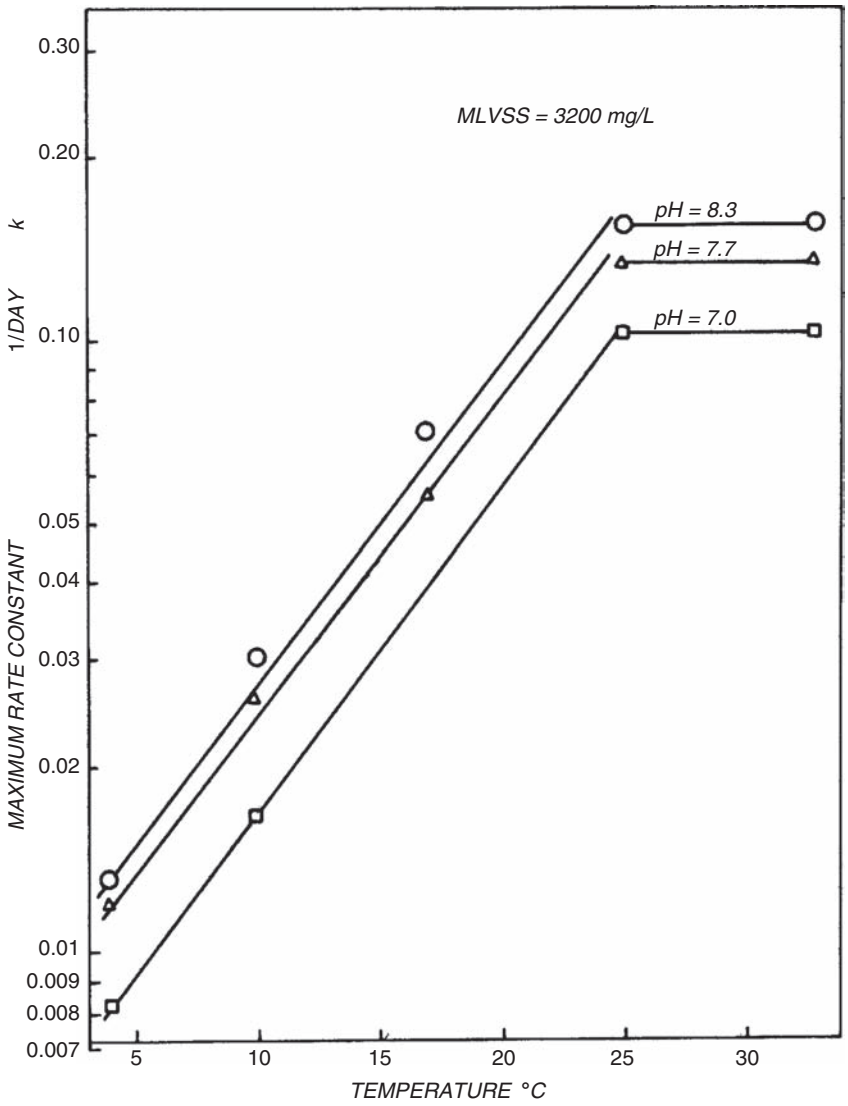

Fig. 1.7. Variation of rate constant with temperature at different pH values (44).

plot reveals that at low temperatures the maximum nitrification velocity does not increase in proportion to the increase in MLVSS. In fact, at 4°C there is very little difference in velocity for the MLVSS concentrations of 430, 1200, and 3200 mg/L. However, the effect of MLVSS on V_m becomes more apparent with increasing temperature. Ultimately, at 25°C and 33°C, V_m is perfectly proportional with MLVSS.

This limitation on V_m caused by low temperatures is shown in Fig. 1.10. The V_m values at different temperatures were plotted against the MLVSS concentrations. The linear plots at 25°C and 33°C reflect the proportionality between V_m and MLVSS. The slope of the lines

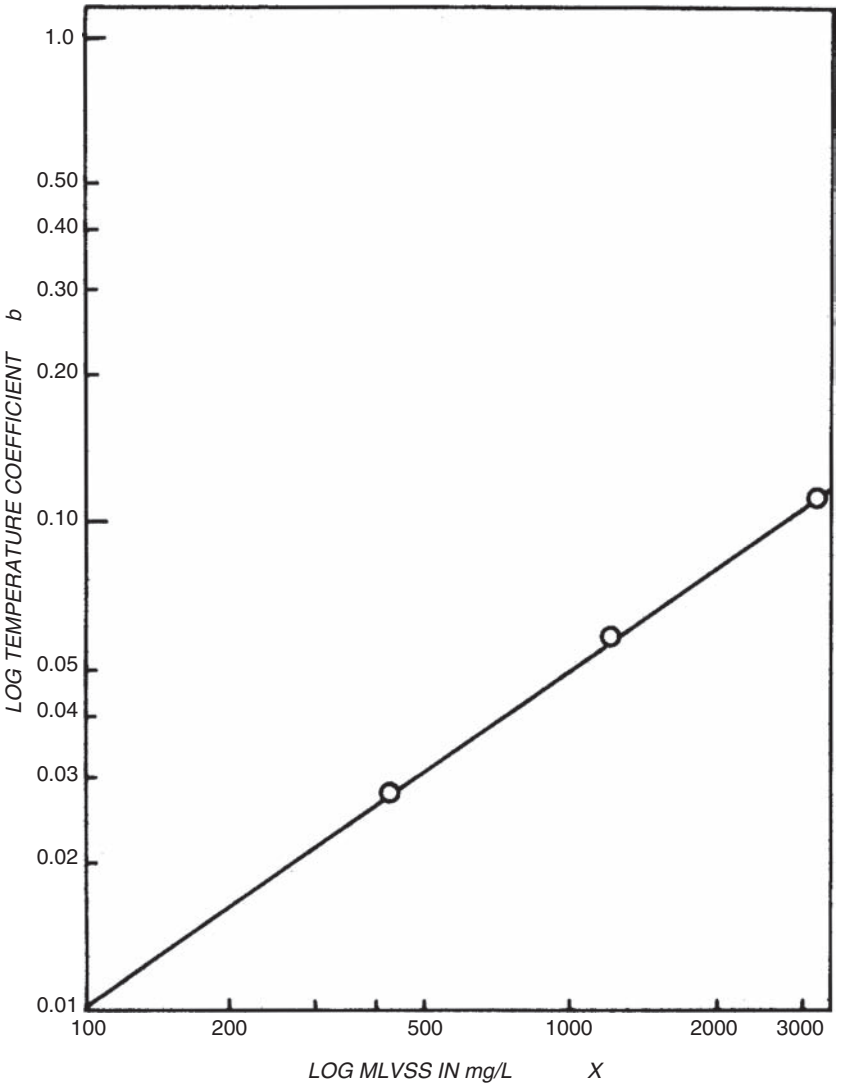

Fig. 1.8. Variation of temperature coefficient with MLVSS concentration (44).

that represent the values of k is constant; however, this does not occur at low temperatures and especially at 4°C, where an increase from 1,200 to 3,200 mg/L MLVSS did not produce any increase in V_m. Subsequently, the calculated value of k, obtained by dividing a constant V_m value by an increased amount of MLVSS, will show a decreasing value for k at increased MLVSS and lower temperatures (Figs. 1.5 to 1.7).

Figure 1.10 suggests that the limitation imposed on the maximum nitrification velocity by temperature is a genuine behavior and not a consequence of variations in experimental procedure. This interpretation is supported by the following observations (44):

Table 1.2
Comparison of temperature effects on maximum nitrification rate constant, *k* (44)

Temperature coefficient (*b*)	Conditions
	Activated Sludge (44)
	Ammonia to nitrate, pH = 7.0 to 8.3
0.028	MLVSS = 430 mg/L, $T = 4°$ to 33°C
0.059	MLVSS = 1,200 mg/L, $T = 4°$ to 25°C
0.121	MLVSS = 3,200 mg/L, $T = 4°$ to 25°C
	Pure culture (52)
0.073	Ammonia to nitrite
	Thames estuary water (46)
0.095	Ammonia to nitrite
0.059	Nitrite to nitrate
	River water (47)
0.084	Ammonia to nitrite
0.056	Nitrite to nitrate
	Activated sludge (51)
0.120	Ammonia to nitrite
	Single stage activated sludge (33, 35) nitrogen removal system
0.075	Nitrification

1. The data for pH 7.0, 7.7, and 8.3 show the same behavior.
2. The activated sludge used in the oxidation rate studies was taken from the same batch of nitrifying sludge.
3. At high temperatures, where the same procedure was used, k was independent of MLVSS concentration.
4. Further elaboration on this behavior will be detailed and modeled in the following section on kinetics.

This study related to the influence of biomass, temperature, and pH on the nitrification rate carried out by Shammas (44) can be summarized in the following eight points:

1. Differing environmental and operating conditions can affect the performance of the nitrification process.
2. There is no interaction between pH and temperature in their effect on the nitrification rate.
3. MLVSS concentration influences the extent of temperature and pH effects so that there is a significant interaction between MLVSS and the other two variables. Consequently, a relationship that expresses the temperature coefficient as a function of MLVSS concentration was developed.
4. The values of the nitrification rate constant k, ranged from 0.0085/d at 4°C and pH 7 to 0.175/d at 33°C and pH 8.3.
5. The modified Arrhenius relationship could be used successfully to estimate the change in nitrification rate with temperature.

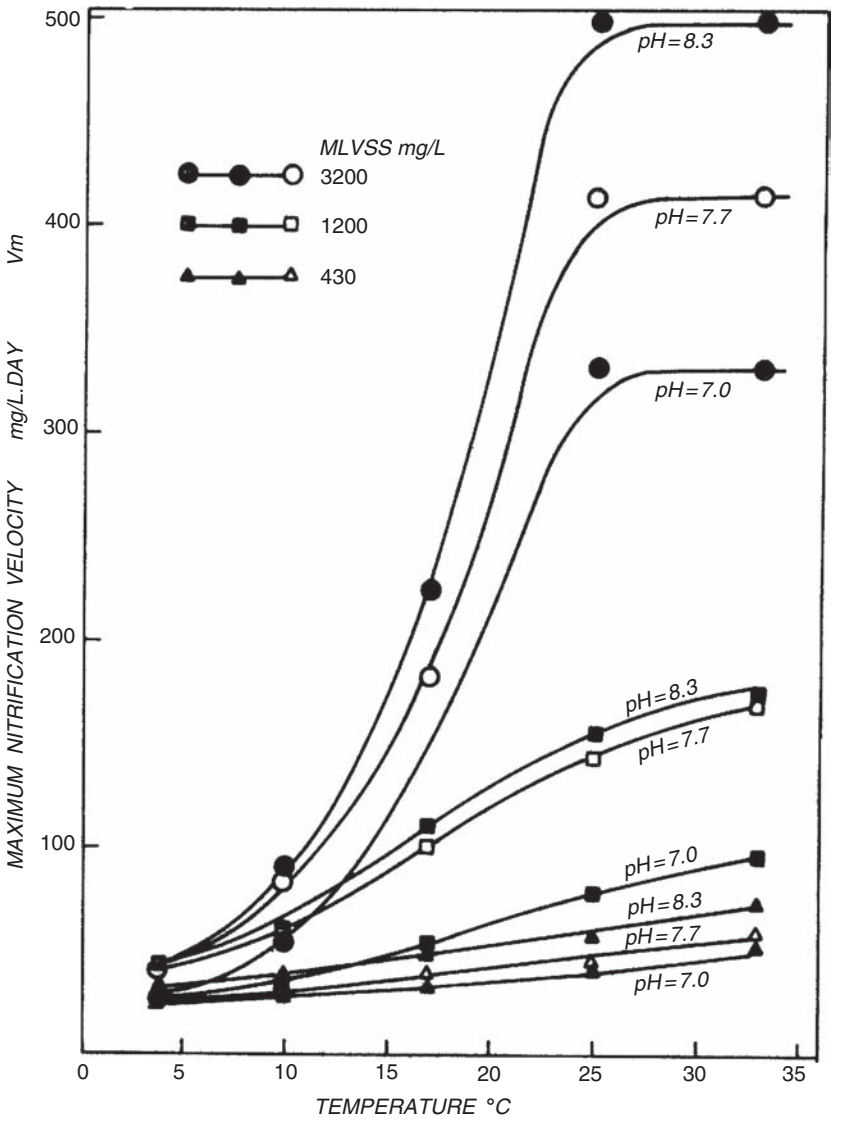

Fig. 1.9. Variation of maximum nitrification velocity with temperature, pH, and MLVSS concentration (44).

6. The extremely depressed nitrification rates at low ammonia concentrations indicate that high nitrification efficiencies can only be obtained with either an unreasonably long detention time or a combination of high mixed-liquor volatile solids concentration and elevated temperature.
7. An increase in the MLVSS concentration at very low temperatures does not significantly improve nitrification efficiency.
8. Design and operation of the nitrification process must be based on combined environmental and operational conditions (pH, temperature, and MLVSS concentration).

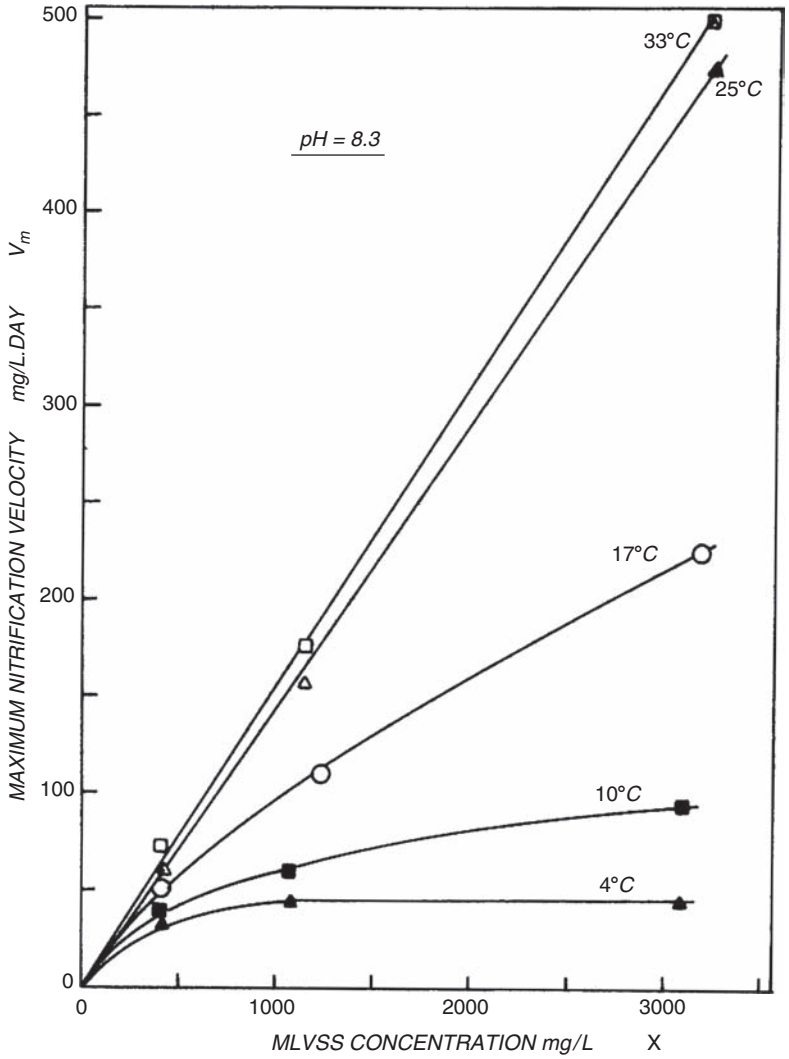

Fig. 1.10. Variation of maximum nitrification velocity with MLVSS concentration at different temperatures (44).

4.3. Design Criteria of Nitrification Systems

This section discusses the design criteria suggested by US EPA (53) as a basis for the design of nitrification suspended growth systems.

4.3.1. Aeration Tank Layout

The tank configuration should insure that flow through the tank follows the plug-flow mixing model as closely as possible. Such configuration can be accomplished by dividing the tank into a series of compartments with ports between them. Figure 1.11 shows three compartments

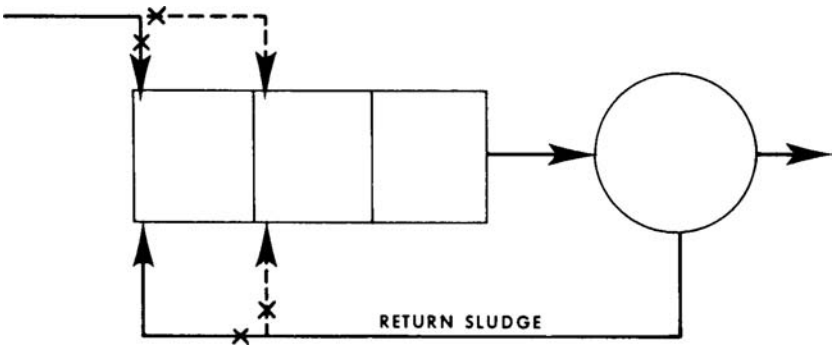

Fig. 1.11. Model nitrification system (53).

as a minimum number. Tanks can be designed for either diffused-air or mechanical-aeration systems. Because the oxidation rate of the process varies widely with temperature, special provisions may be necessary to incorporate the necessary flexibility in the oxygen supply system.

4.3.2. pH Control

Nitrification tanks should be sized to permit complete nitrification under the most adverse combination of ammonia load and temperature expected, and at a pH as near optimum as feasible. The range of 7.6 to 7.8 is recommended to allow carbon dioxide to escape to the atmosphere.

The nitrification process destroys alkalinity and the pH may fall to concentrations that will inhibit nitrification unless excess alkalinity is present in the wastewater or lime is added to maintain favorable pH concentrations.

$$2NH_4HCO_3 + 4O_2 \rightarrow 2HNO_3 + 4H_2O + 2CO_2 \quad (55)$$

$$2HNO_3 + Ca(HCO_3)_2 \rightarrow Ca(NO_3)_2 + 2CO_2 + 2H_2O \quad (56)$$

Overall, the addition of Equations (55) and (56) yields:

$$2NH_4HCO_3 + 4O_2 + Ca(HCO_3)_2 \rightarrow Ca(NO_3)_2 + 4CO_2 + 6H_2O \quad (57)$$

Theoretically, 7.2 lb of total alkalinity are destroyed per pound of ammonia nitrogen oxidized to nitrate. One-half of this destruction is attributable to loss of alkalinity caused by ammonia and the remainder is attributable to destruction of natural alkalinity, as shown in Eqs. (55) to (57). Whether lime additions will be required depends upon the alkalinity of the wastewater and the desired pH of operation. For operation under the most adverse temperature conditions and at operating pH, enough lime must be added initially to raise the pH into the desired range, and then 5.4 lb of hydrated lime per pound of ammonia nitrogen will be required to maintain the pH. An actual titration test should be conducted to obtain design criteria. In Boston sewage, about 250 lb of hydrated lime are needed per MG to raise the pH initially to optimum pH range, and an additional 700 lb are needed to hold it there during the course of oxidation of the ammonia. The total hydrated lime requirements are estimated to be about

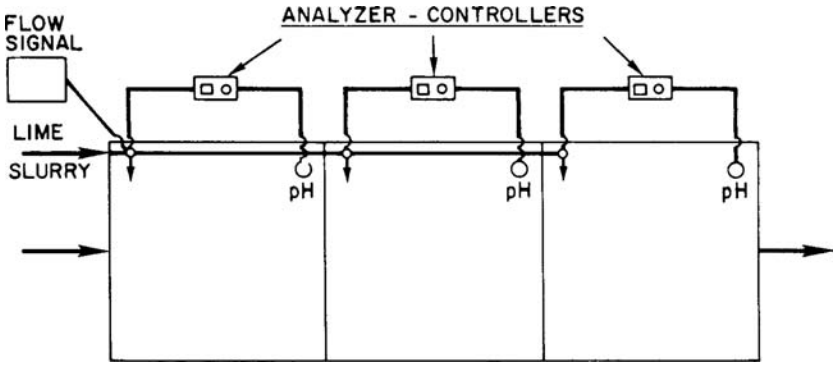

Fig. 1.12. pH control for nitrification system, plan view (53).

115 mg/L. Additional amounts of lime may be required if chemicals, such as alum, have been added previously for phosphorus removal.

Marked reductions in lime requirements will result in any system that can be designed to operate at pH levels of 7.8 or less, because carbon dioxide resulting from destruction of alkalinity and organic matter will be washed out of the liquid phase by air contact. The pH of such systems will vary with the rate of aeration (ventilation). The type and sensitivity of the pH control system will depend on the character of the wastewater and the variations in the ammonia load fed to the system. Fig. 1.12 shows a proposed system for pH control in the most demanding situation. In many situations, a lesser degree of control will be feasible; in some none will be needed.

In any event, enough alkalinity should be present to leave a residual of from 30 to 50 mg/L after nitrification is completed. As a general rule, where phosphorus removal is accomplished in the first stage of a two- or three-stage system by use of alum or ferric salts, it will be necessary to provide lime-feeding facilities, and the optimum pH of operation becomes more or less an academic matter. In situations where feeding of lime is not essential, good engineering normally will indicate that additional tankage be provided to overcome the limitations of reduced activity, as opposed to providing lime-feeding facilities to keep the tankage at a minimum.

4.3.3. MLSS and MLVSS Concentrations

Designs based on MLSS (mixed liquor suspended solids) concentration alone should be avoided, because MLSS will not truly reflect the biological mass in the system. The ratio of MLVSS (mixed liquor volatile suspended solids) to MLSS may vary depending on the nonvolatile suspended solids (including residual chemical precipitates) in the feed. For nitrification systems receiving normal secondary effluents, MLVSS concentrations of 1,500 to 2,000 mg/L seem to be safe for design.

4.3.4. Aeration Tank Capacity

The choice of the design-peak load depends on the circumstances of the specific project, and need not necessarily be the absolute maximum expected load. For many projects, the

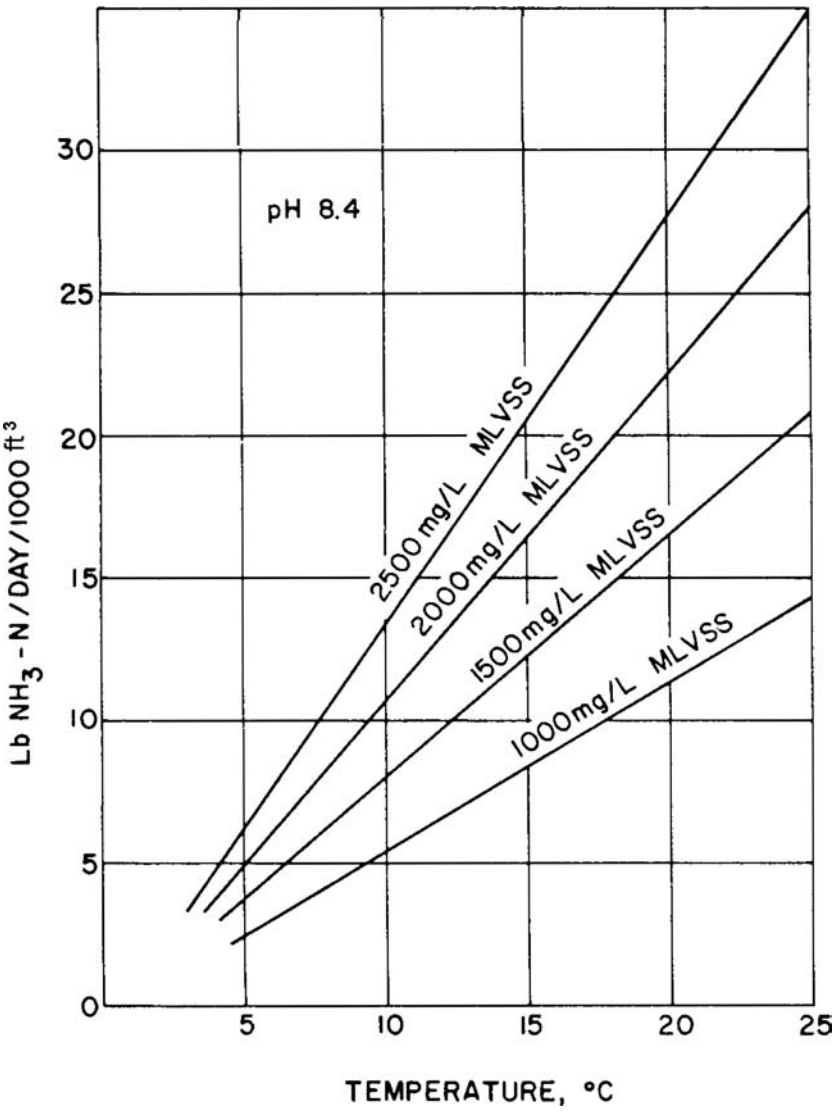

Fig. 1.13. Permissible nitrification-tank loadings (53).

use of a peak-load factor of 1.5 represents a reasonable peak at low-temperature conditions. Fig. 1.13 shows the permissible volumetric loading of the nitrification tanks at a pH of 8.4 and at various temperatures and MLVSS concentrations.

Figure 1.14 shows the corrections that must be applied to the permissible loadings when the pH is different from 8.4. In plants with well-buffered wastewater, it may be more economical to provide the additional tankage to permit operation at a lower pH, rather than to add an alkaline material.

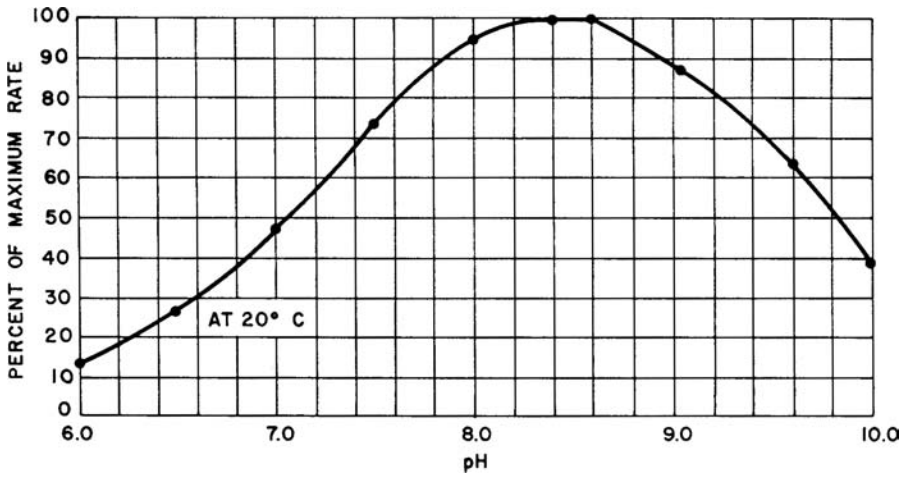

Fig. 1.14. Percent of maximum rate of nitrification at constant temperature versus pH (53).

4.3.5. Oxygen Requirements

Stoichiometrically, each pound of ammonia nitrogen that is nitrified requires 4.6 lb of oxygen. (The amount of ammonia nitrified is usually slightly more than the amount of nitrate measured because some denitrification occurs.) Usually, it is assumed that all of the ammonia fed will be nitrified. An additional oxygen allowance must be made for carbonaceous BOD that escapes from the secondary treatment process.

Nitrification seems to be uninhibited at DO concentrations of 1 mg/L or more. Design based on maintaining 3 mg/L of DO in the mixed liquor under average loading conditions includes a reasonable factor of safety. Under peak loading the DO concentration may be permitted to fall, but not below 1 mg/L.

The rate of nitrification will vary significantly with temperature and pH, and compensation for this variation must be made in the design of the plant. During the summer, the following methods can be used to match the oxygen demand rate to the oxygen supply rate:

1. Reduce MLSS concentration
2. Reduce pH by reducing chemical supply
3. Reduce tankage in service while increasing oxygen supply to the tanks remaining in service

4.3.6. Settling Tanks

Surface Loadings: The maximum permissible hydraulic surface loading is $1,000\ \text{gpd/ft}^2$. Average surface loadings should be in the range of 400 to $600\ \text{gpd/ft}^2$. It may be necessary to reduce this loading if the MLSS concentration is greater than 2,500 mg/L, because of limiting sedimentation-tank-solids loadings.

Number of Tanks: Because of the relatively slow growth and settling rates of nitrifying biosolids it is desirable to provide more than two settling tanks to insure that the biosolids are kept within the system when a tank is down for maintenance and repair. Four tanks are a desirable minimum number.

Depth: Depths of 12 to 15 ft are recommended.

Biosolids-Collection Equipment: Experience has shown no evidence of rising biosolid problems, probably attributable to complete nitrification and very low residual carbonaceous BOD concentrations. Use of rapid-removal suction-type biosolids-collection equipment is not mandatory, but it may be desirable in large circular tanks. The settling tanks should be equipped with skimmers and provision should be made to use the scum system to pump floating biosolids, should it ever occur, to the nitrification tank influent.

Biosolids: It is recommended that capacity be provided for a return-biosolids rate of 50% to 100% of average flow, because the nitrification biosolids are lighter and do not compact as well as carbonaceous biosolids. Continuous biosolids wasting is not normally necessary. Periodic adjustments of MLSS concentration are necessary, however, and provisions should be made to dispose of waste nitrification biosolids with the waste biosolids from the carbonaceous treatment process.

5. KINETICS OF THE NITRIFICATION PROCESS

As was discussed earlier in Section 4 various reaction rate equations have been suggested in the literature to describe the nitrification process. Shammas (29, 44, 54) carried out an extensive and systematic kinetic analyses under various controlled operational conditions to determine the best suitable kinetic model.

The sigmoidal characteristic exhibited in the variation of the reaction velocity as a function of ammonia-N concentration led to the adoption for the first time of an allosteric kinetic model in the environmental engineering field (54). The nitrification data obtained were found to have an excellent fit to the model. The model parameters were determined, thus making it possible to predict the extent of nitrification under a given set of operational and environmental conditions. The results and conclusions of this study would help in the process of obtaining an optimum design and operation of a wastewater treatment plant that is designed and operated for the extraction of nitrogen from its influent.

5.1. Analysis of Nitrification Data

A typical plot of reaction velocity against ammonia nitrogen concentration is shown in Fig. 1.15. All data at different concentrations of MLVSS, pH, and temperature showed the same form of curves indicating a sigmoidal characteristic. This distribution cannot be fully described by the hyperbolic kinetic model of Monod. This fact is made clearer by plotting the inverse of the Monod relationship as shown in Eq. (51):

$$\frac{1}{v} = \frac{1}{\mathrm{d}S/\mathrm{d}t} = \frac{1}{kX} + \frac{K_\mathrm{s}}{kX}\frac{1}{S} = \frac{1}{V_\mathrm{m}} + \frac{K_\mathrm{s}}{V_\mathrm{m}}\frac{1}{S} \tag{51}$$

The plot of $1/(\mathrm{d}S/\mathrm{d}t)$ versus $1/S$ should be a straight line. However, the plot of the calculated data, as illustrated in Fig. 1.16, indicates otherwise. At high substrate concentrations, the points do fit a straight line relationship, while at low ammonia-nitrogen concentrations, the

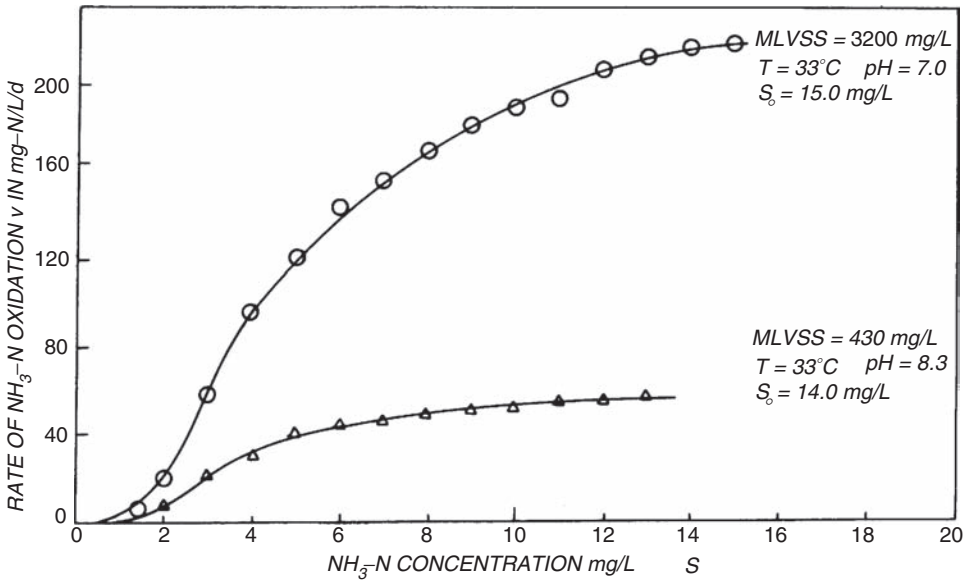

Fig. 1.15. Variation of nitrification rate with NH_3 concentration (29).

curves are far from being linear. Evidently, a different and more complicated kinetic model is needed to describe the sigmoidal character in the variation of reaction velocity as a function of ammonia nitrogen concentration (54).

5.2. Allosteric Kinetic Model

It has been shown that many enzymes portray kinetic properties in such a way that the velocity as a function of substrate concentration shows a sigmoidal dependence. This realization had led to the proposal of several theories to explain this phenomenon (55–60). Of these, the one that has received experimental support and gained the widest acceptance (61, 62) is the allosteric model of Monod, Wyman, and Changeux (55).

The Monod, Wyman, and Changeux model (M–W–C model) comprises both new ideas and terminology that have not been touched upon in the water pollution and waste treatment literature. Therefore it seems appropriate at this point to introduce the principles upon which Monod and co-workers based their model (Fig. 1.17):

1. Allosteric enzymes are polymers that are composed of identical and finite number of subunits (protomers).
2. Allosteric enzymes are proteins having several substrate binding sites. (Fig. 1.17 illustrates a dimer, i.e., number of binding sites, n, equals 2.).
3. Allosteric effects result from the interaction among such distinct specific sites.
4. These allosteric effects are brought about by a molecular transition (allosteric transition), which is induced in the enzyme after binding with the substrate (allosteric ligand).
5. Two states are reversibly accessible to allosteric polymers. These states (R and T) differ by the distribution and/or energy of interprotomer bonds.

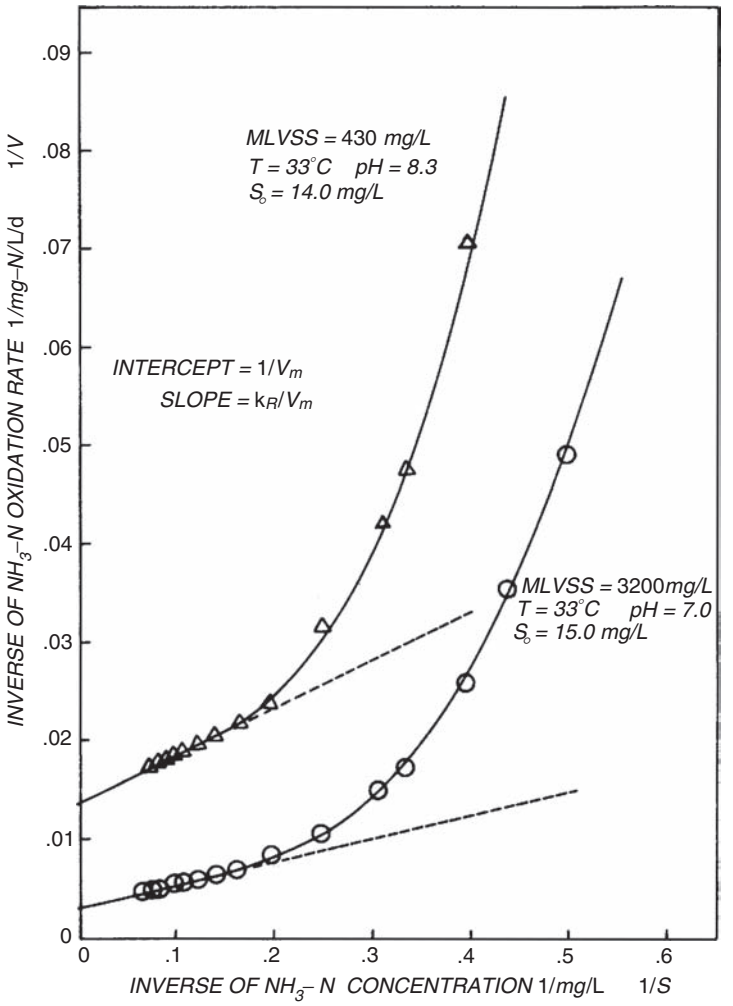

Fig. 1.16. Plot for specific rate determination (29).

6. As a result, the affinity of one (or several) of the stereo-specific sites towards the corresponding ligand is altered when a transition occurs from one to the other state (L is the equilibrium constant for the $R_0 \leftrightarrow T_0$ transition).

Based on the equilibria expressions for the R and T states and considering the probability for the dissociation of the different complexes Monod and coworkers (55) derived the following M–W–C saturation function:

$$Y = \frac{Z(1+Z)^{n-1} + LcZ(1+cZ)^{n-1}}{(1+Z)^n + L(1+cZ)^n} \tag{58}$$

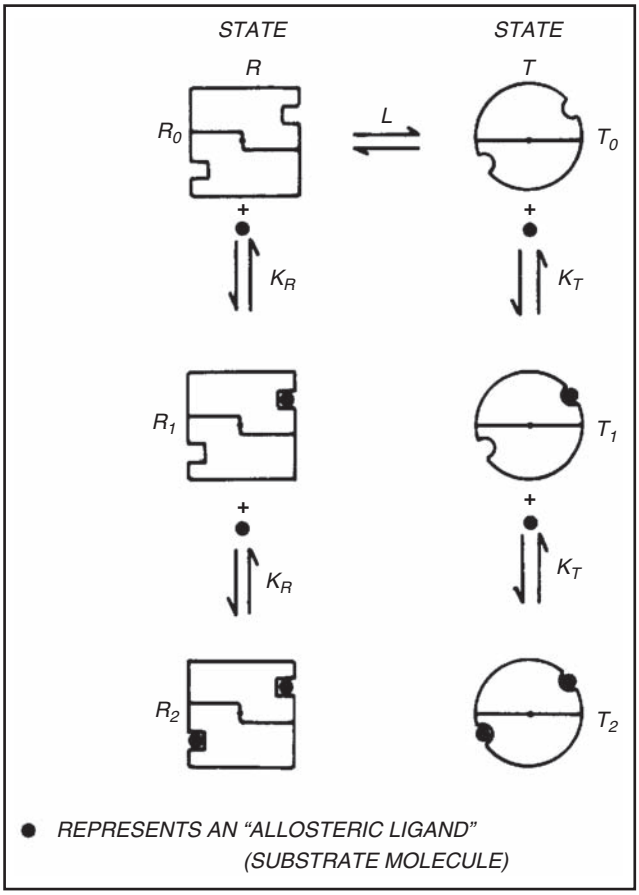

Fig. 1.17. Schematic diagram of M–W–C model illustration for an enzyme having two binding sites (29).

where,

Y = the fraction of sites bound with substrate
= saturation function, a dimensionless number

Z = the reduced substrate concentration, a dimensionless number = S/K_{R}

K_{R} = the microscopic dissociation constants of a substrate (S) bound to a site in the R state

K_{T} = the microscopic dissociation constants of a substrate (S) bound to a site in the T states

n = number of binding sites or protomers
= an interaction coefficient

S = substrate concentration

c = a dimensionless number, $c = K_{\mathrm{R}}/K_{\mathrm{T}}$

L = allosteric constant. It represents the equilibrium constant for the $R_0 \leftrightarrow T_0$ transition (i.e., in the absence of substrate)
T, R = represent the two states or conformations of the enzyme

5.3. *Application of M–W–C Model to Nitrification*

The velocity of a biochemical reaction is proportional to the sites actually bound by the substrate. In the case of simple enzyme kinetics, the velocity of the reaction (v) is proportional to the concentration of the enzyme–substrate complex, so that:

$$v = k[\mathrm{ES}] \tag{59}$$

where
k = specific rate constant or max rate of NH_3-N utilization per unit weight of microorganisms
mg NH_3-N/mg MLVSS/d or 1/d
v = rate of substrate utilization, mg/L NH_3-N/d or mg/L/d
[ES] = concentration of the enzyme–substrate complex, mg/L
In the allosteric case, the velocity of the reaction is given by (63),

$$v = kEY \tag{60}$$

where
E = total enzyme concentration, mg/L

Thus the M–W–C equation can be expressed as follows:

$$v = kE\frac{Z(1+Z)^{n-1} + LcZ(1+cZ)^{n-1}}{(1+Z)^n + L(1+cZ)^n} \tag{61}$$

It is of interest to note that when the allosteric constant, L, approaches zero, or in other words, for enzymes showing no allosteric effects, the M–W–C kinetic equation simplifies to:

$$v = kE\frac{Z(1+Z)^{n-1}}{(1+Z)^n} \tag{62}$$

$$v = kE\frac{Z}{1+Z} \tag{63}$$

and because $Z = S/K_R$, therefore:

$$v = \frac{kES}{S + K_R} \tag{64}$$

which is the known and familiar Michaelis–Menten kinetic model (54).

At this point a comparison between Michaelis–Menten expression and the Monod model, Eq. (50), introduced previously in the form:

$$v = \frac{kXS}{S + K_s} \tag{50}$$

is in order. It is quite clear that both expressions have the same form (note that when there is one state of the enzyme then $K_S = K_R$), although the former was based on theoretical analysis, while the latter was purely empirical, and based on substrate utilization studies in biological processes (54). In the first expression the velocity is proportional to the enzyme concentration, while in the second, it is proportional to the microorganisms' concentration. By applying the same analogy to the M–W–C model, one can rewrite the kinetic expression in the following final form (54):

$$-\frac{dS}{dt} = v = kX\frac{Z(1+Z)^{n-1} + LcZ(1+cZ)^{n-1}}{(1+Z)^n + L(1+cZ)^n} \tag{65}$$

or

$$-\frac{dS}{dt} = v = V_m\frac{Z(1+Z)^{n-1} + LcZ(1+cZ)^{n-1}}{(1+Z)^n + L(1+cZ)^n} \tag{66}$$

5.4. Determination of Kinetic Parameters

The procedure for the analysis of experimental data for the determination of the kinetic model parameters is based on the use of the limiting conditions along with the general features of the allosteric kinetic model. The mode of analysis and basic interpretations were developed by Frieden (56) and by Boring and Horon (64). For details of the procedure, the reader is referred to the above two references.

A complete listing of the values of parameters obtained by using the above technique (54), for various concentrations of MLVSS, pH, and temperature values, is shown in Tables 1.3 to 1.5.

Figure. 1.18 illustrates a typical fit of curves for a MLVSS concentration of 3,200 mg/L and a pH value of 7 (54). The curves shown are for temperature values of 4°C, 10°C, 17°C, 25°C, and 33°C. Figure. 1.19 shows these same data plotted in terms of Y against Z. This is comparable to a plot showing the variation of reaction velocity with substrate concentration. The deviation of the kinetic data from the hyperbolic function (Michaelis–Menten or Monod) is quite evident. The deviation seems to widen as the temperature increases from 4°C to 10°C (this is reflected in the increase of the L value). The gap narrows at 17°C and reaches a minimum at 25°C and 33°C. Generally the same behavior was found for other pH and MLVSS values (54).

The excellent fit of the M–W–C model to experimental data is illustrated in Figs. 1.18 and 1.19. The same condition was obtained at all other environmental conditions. This indicates that the allosteric model could be an extremely useful tool for the analysis and prediction of ammonia oxidation rates under different environmental conditions (54). It should be remembered that fit alone is no absolute proof of allosterism. This may be only proved by isolating the specific enzymes and running binding studies, which is a rather complicated and uncertain procedure. It has been found that most allosteric enzymes dissociate and lose their allosteric properties on isolation and purification (62). However, the lack of enzyme verification does not decrease the importance and value of the theoretical model. As elucidated below, the model describes quantitatively and in detail the complex interrelationships between the parameters involved and their overall effect on the nitrification rate (54).

Table 1.3
Variations of M–W–C model parameters with pH and temperature-MLVSS = 430 mg/L (29)

pH	$T(°C)$	MLVSS (mg/L)	S_0 (mg/L)	V_m (mg/L/day)	k (mg N/mg MLVSS/day)	K_R (mg/L)	c	n	L
7.0	4	470	13.5	24.9	0.0530	22	0	15	2,000
	10	470	13.5	29.9	0.0635	19	0	15	2,000
	17	458	14.0	35.7	0.0780	16	0	11	100
	25	430	14.1	41.7	0.0970	5.7	0	6	100
	33	430	14.1	53.2	0.124	4.7	0	6	100
7.7	4	436	13.9	27.5	0.0630	20	0	15	200
	10	436	13.9	32.7	0.0750	10	0	6	20
	17	419	13.9	40.0	0.0955	8.8	0	6	50
	25	421	14.5	45.5	0.108	5.6	0	6	20
	33	421	14.5	58.8	0.140	3.8	0	6	2
8.3	4	436	12.7	34.3	0.0785	14	0	15	1000
	10	436	12.7	40.0	0.0916	7.2	0	15	50,000
	17	430	14.2	50.0	0.116	5.4	0	11	5,000
	25	420	14.0	55.6	0.132	4.4	0	6	50
	33	420	14.0	73.5	0.175	3.5	0	6	25

Table 1.4
Variations of M–W–C model parameters with pH and temperature-MLVSS = 1200 mg/L (29)

pH	T(°C)	MLVSS (mg/L)	S_0 (mg/L)	V_m (mg/L/day)	k (mg N/mg MLVSS/day)	K_R (mg/L)	c	n	L
7.0	4	1360	13.6	26.3	0.0193	20	0	15	500
	10	1360	13.6	35.7	0.0262	15	0	15	500
	17	1140	14.3	51.3	0.0450	1.5	0	15	–
	25	1186	13.9	79.4	0.0670	3.4	0	15	50
	33	1186	13.9	96.2	0.0812	4.0	0	15	50
7.7	4	1200	10.6	44.3	0.0370	19	0	15	300
	10	1200	10.6	62.5	0.0520	8.8	0	15	15,000
	17	1250	14.2	103	0.0825	4.3	0	15	3,000
	25	1136	15.2	143	0.126	5.7	0	15	1,000
	33	1136	15.2	169	0.149	6.8	0	15	20
8.3	4	1086	8.3	42.3	0.0390	13	0	15	300
	10	1086	8.3	58.8	0.0540	7.0	0	15	3,000
	17	1252	13.2	110	0.0877	4.8	0	15	1,000
	25	1160	11.8	154	0.133	6.2	0	15	10
	33	1160	11.8	175	0.151	7.0	0	15	10

Table 1.5
Variations of M–W–C model parameters with pH and temperature-MLVSS = 3200 mg/L (29)

pH	T(°C)	MLVSS (mg/L)	S_0 (mg/L)	V_m (mg/L/day)	k (mg N/mg MLVSS/day)	K_R (mg/L)	c	n	L
7.0	4	3,260	15.6	27.8	0.0085	9.5	0	15	10,000
	10	3,260	15.6	54.9	0.0169	3.8	0	15	5×10^6
	17	3,200	15.0	256	0.0800	7.4	0	15	600
	25	3,200	15.0	333	0.104	8.0	0	15	75
	33	3,200	15.0	333	0.104	8.0	0	15	75
7.7	4	3,240	13.3	39.2	0.0121	14	0	15	300
	10	3,240	13.3	84.0	0.0260	3.0	0	15	40,000
	17	3,220	11.4	182	0.0565	2.2	0	15	10×10^6
	25	3,120	13.8	416	0.133	12	0	15	35
	33	3,120	13.8	416	0.133	12	0	15	35
8.3	4	3,100	13.3	40.6	0.0131	7.3	0	15	40,000
	10	3,100	13.3	93.3	0.0301	2.5	0	15	400,000
	17	3,170	14.0	225	0.0710	3.0	0	15	–
	25	3,240	15.2	500	0.154	16	0	15	15
	33	3,240	15.2	500	0.154	16	0	15	15

5.4.1. Dissociation Constants (K_R, K_T) *and* c

The ratio of dissociation constants ($c = K_R/K_T$) was always zero, for all values of microbial mass, pH, and temperature that were investigated. This indicates that there is an exclusive substrate binding by one conformation of the enzyme (54). For this condition of $c = 0$, the kinetic equation reduces to:

$$Y = \frac{Z(1+Z)^{n-1}}{L + (1+Z)^n} \tag{67}$$

because $Y = v/V_m$, $v = -dS/dt$ and $Z = S/K_R$, the nitrification rate equation can be expressed as:

$$-\frac{dS}{dt} = V_m \frac{\frac{S}{K_R}\left(1 + \frac{S}{K_R}\right)^{n-1}}{L + \left(1 + \frac{S}{K_R}\right)^n} \tag{68}$$

It should be noted that for $c = 1$, that is when the affinity of both conformations towards the substrate is the same, the saturation function simplifies to

$$Y = \frac{Z}{1+Z} = \frac{S}{S + K_R} \tag{69}$$

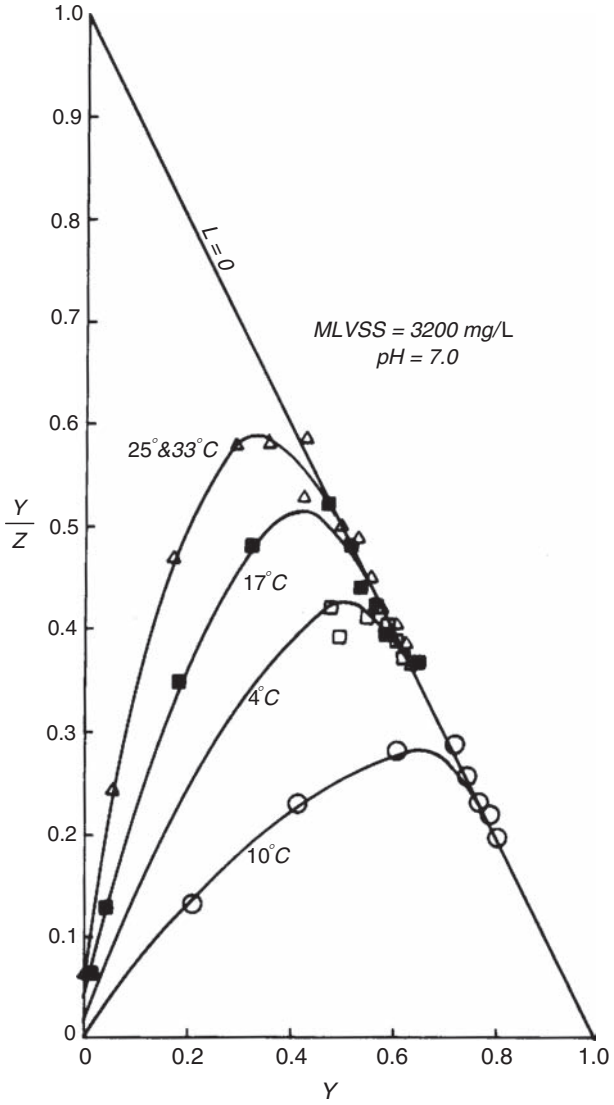

Fig. 1.18. Fit of data to theoretical curve (29).

or,

$$v = kEY = \frac{kES}{S + K_R} \tag{70}$$

which is the Michaelis–Menten model, Eq. (50). Thus, it is clear that with respect to the value of c, the data belong to a set having the maximum deviation from the hyperbolic function that this model provides. In other words for $c = 0$, the sigmoid characteristic of the data is a

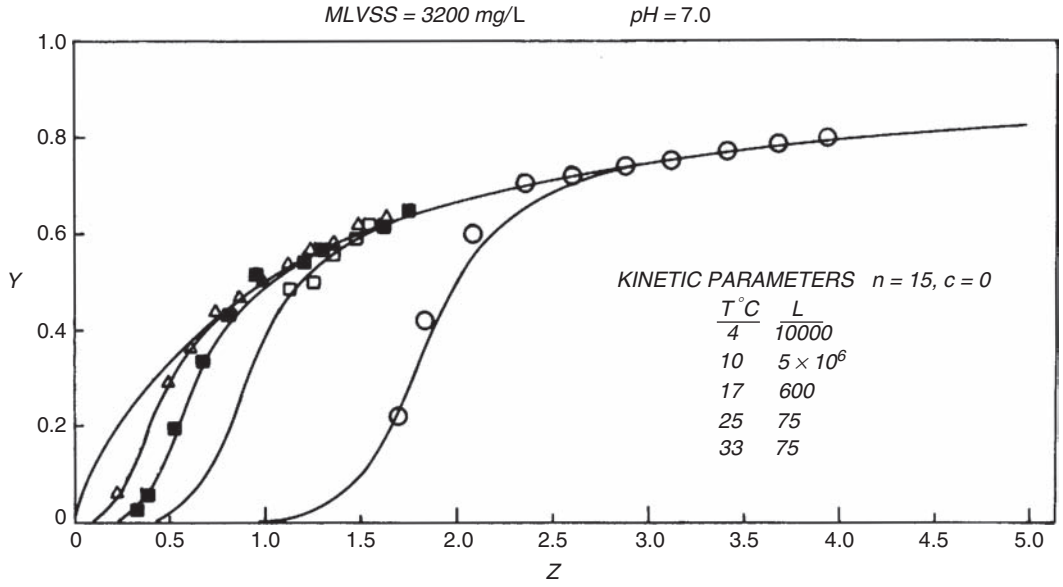

Fig. 1.19. Fit of data to theoretical curve (29).

maximum (considering that L is constant). However, it should be pointed out that at low values of the allosteric constant, L, and whatever the value of c may be, the sigmoid characteristic is minimal (54).

A consequence of the fact that $c = 0$, is that the dissociation constant K_T of the substrate and T state of the enzymes is infinite. As a result the T conformation could exist only in the unbound state T_0. The binding of the substrate to the R conformation is reflected in the dissociation constant K_R. The variation of this parameter with temperature and pH at different MLVSS concentrations is shown in Table. 1.3 to 1.5. At low MLVSS (430 mg/L), Table 1.3 indicates that the dissociation constant decreases with increasing temperature (4°C to 33°C), and pH (7.0 to 8.3). It should be pointed out that the velocity of ammonia utilization is inversely proportional to the K_R value. Consequently the decrease in the value of K_R with increasing temperature and pH indicates an increase in the velocity of nitrification. However, for the higher concentration of MLVSS at 1,200 mg/L (Table 1.4), although the K_R value does decrease with increasing temperature and pH up to 17°C, a reverse trend occurs at the higher temperature. This reversal in relationship seems to occur in the temperature range of 14°C to 20°C. A very small variation in the value of K_R occurs after 25°C.

The interesting feature of this change in behavior created by the increase in microbial mass is that it altered both the pH and the temperature effects as far as K_R is concerned. It should also be noticed that the shift in trend is much more pronounced at the highest MLVSS concentration. For a microbial mass of 3,200 mg/L the values of K_R at 25°C and 33°C were far higher than those at 10°C to 17°C (54). The minimum values of K_R (which correspond

to higher oxidation velocities) at MLVSS of 430, 1,200, and 3,200 mg/L occur at temperature values of 25°C to 33°C, 17°C, and 10°C to 17°C, respectively for all pH values.

The minimum value of K_R at MLVSS of 430 mg/L occurs at a pH value of 8.3 for all temperatures, whereas at the higher concentrations of MLVSS, the minimum value of K_R is at either pH 8.3 or 7.0 depending on whether the temperature is below or above 17°C.

It is of special interest that a significant interaction among the environmental parameters exists. In other words, each of the three parameters studied is not affecting the dissociation constant independently. Also it is unrealistic to compare the value of K_R with K_m of the hyperbolic function. For, in addition to the basic difference in their significance, it can be shown that to some extent the value of K_R is related to K_m by the following expression (64):

$$K_R = f(K_m)^{1/n} \tag{71}$$

Thus any variation in the value of the interaction coefficient, n, (or number of binding sites) by temperature, pH, or microbial mass will be reflected in the value of K_R. No studies are available for comparison. Kirschner (65) states that many challenging questions remain open. Among the first is the pH and temperature dependence of the model parameters.

5.4.2. Allosteric Constant, L

One of the main assumptions of the model is that allosteric effects "are attributable to the displacement of equilibrium among discrete states assumed to exist, at least potentially, apart from the binding of a substrate" (55). Thus the allosteric constant, L, defines "the contribution of the protein itself to the interaction, as distinct from the dissociation constants of the substrates." Because L represents an equilibrium constant, it would be expected to show a temperature and pH dependence. Examination of Table. 1.3 to 1.5 indicates that far more allosteric effect is exhibited at lower temperatures. The ratio of the allosteric constant at 25°C to that at 4°C varied from 10 to 4,000 times. At higher MLVSS, even greater ratios were shown between those at 25°C and 10°C. As far as the pH is concerned, the effect was much less dramatic than that of temperature. In general, the allosteric constant was smaller at pH values of 7.7 and 8.3 than at pH 7.0 (54).

The significance of these results on the nitrification process should be made clear. The degree of sigmoid characteristic in the data is, to a great extent, represented by the allosteric constant, L. Furthermore, allosterism, or sigmoidness is more exhibited, the lower the substrate concentration. Also the lower the substrate concentration, the smaller becomes the saturation function Y. Consequently, the ratio of v to V_m gets smaller with increasing values of L; as a result, a very low percentage of the maximum possible velocity is attained at low temperatures and substrate concentrations. When this effect is combined with the slow reaction rate displayed at depressed temperatures (*see* Section on "*k*"), extremely low efficiencies will be shown in the nitrification process (54). Based on these findings, it is recommended to treat the ammonia-rich digester supernatant in combination with the raw wastewater. If the temperature is low, then heating the return sludge becomes an attractive possibility. However, in case it is deemed necessary to remove the last mg/L of ammonia nitrogen from the treated effluent, it is suggested that a supplemental process be used. Breakpoint chlorination is a possibility (54).

5.4.3. Number of Binding Sites, n

Monod and coworkers (55) state that oligomeric enzymes undergo conformational changes and dissociate into smaller oligomers, and that this dissociation is markedly dependent on temperature. Examination of Table 1.3 indicates a similar behavior. The number of sites *n* decreased from 15 at 4°C to 6 at 25°C and 33°C. The same tendency for dissociation was exhibited at the three different pH values. However, for 10°C and 17°C, smaller values of *n* were obtained at pH 7.7 than at either 7.0 or 8.3. This may imply that at certain temperatures some association among oligomers takes place as the pH increases from 7.7 up to 8.3 (54).

Ipata and Cercignani (66), in a study on the enzyme 5′-nucleotidase, reported the same variations in the value of n at pH 7, 7.5, and 8.0. They reasoned that this might be a consequence of a change in the protein structure, causing a variation in the accessibility of the binding sites.

For higher MLVSS (1,200 and 3,200 mg/L) shown in Table. 1.4 and 1.5, no variation in the value of n occurred at any temperature or pH. This tendency by the enzymes to undergo association at high enzyme concentrations and dissociate at low concentrations has been reported for many enzymes, including glutamate dehydrogenase, and phosphorylase a (56). Because of this mechanism of association–dissociation, Frieden (56) explains that an enzyme may show normal kinetic behavior (Michaelis–Menten) at low enzyme concentrations, and a sigmoidal behavior at higher concentrations of enzyme concentration.

Because the kinetic behavior is dependent on the enzyme concentration, it stands to reason that the kinetic properties of such enzymes may depend upon the particular protomers present and their relative concentrations. Such behavior of allosteric enzymes may explain the dependence of the kinetic parameters on the MLVSS concentration as shown in these data (54).

5.4.4. Maximum Nitrification Rate Constant, k

The values of the maximum nitrification rate *k* (Table. 1.3 to 1.5) ranged from a low of 0.0085 at 4°C and pH 7, to a high of 0.175 mg/L NH_3-N/mg/L MLVSS/d at 33°C and pH 8.3. These values show good correlation with the results indicated by different studies on nitrifier enriched activated sludge. However, they are much lower than those reported for either river water or pure culture. In fact this should be expected because in activated sludge, the rate is based on MLVSS, which even when enriched with nitrifiers, is not completely composed of nitrifying organisms (54). Painter and Jones (48) reported that the highest rate they could attain was 0.144, and that the usual rate was between 0.05 and 0.07 mg N/mg MLVSS/d. The maximum rate they obtained was only about 2% of that of a pure culture. Wild, Sawyer, and McMahon (49) found that the rate varied from a maximum 0.185 at a pH of 8.4 to a minimum of 0.020 mg N/mg MLVSS/d at a pH of 6. Bishop et al. (34) reported a rate of 0.11 mg N/mg MLVSS/d at 27°C that decreased down to 0.032/d at 15°C. Sutton et al. (39) showed that at a MLVSS concentration of 1,700 mg/L, pH 7 to 8, and a temperature of 21°C, the rate, k, was 0.0216 mg N/mg MLVSS/day. They also reported that the sludge retention time had to be doubled from 30 up to 60 ds to attain the same extent of nitrification at 10°C.

The effect of temperature and pH on *k* at different MLVSS concentrations indicates that the optimum operating temperature and pH values are just above 25°C and 8.0, respectively (54, 67–72).

The conclusion is that the sigmoidal characteristics exhibited in the variation of the velocity of the nitrification process can be well described by an allosteric kinetic model. The nitrification data were found to have an excellent fit to the model. The model parameters were determined thus making it possible to predict the extent of nitrification under a given set of operational conditions (54).

6. DENITRIFICATION BY SUSPENDED GROWTH SYSTEMS

To achieve the desired nitrogen removal it is required to follow the nitrification process by denitrification to convert the nitrate-nitrogen to nitrogen gas (71–76)

US EPA Report (53) on nitrification and denitrification facilities contains information that may serve as a basis for the design of a denitrifying suspended growth system. Figure 1.20

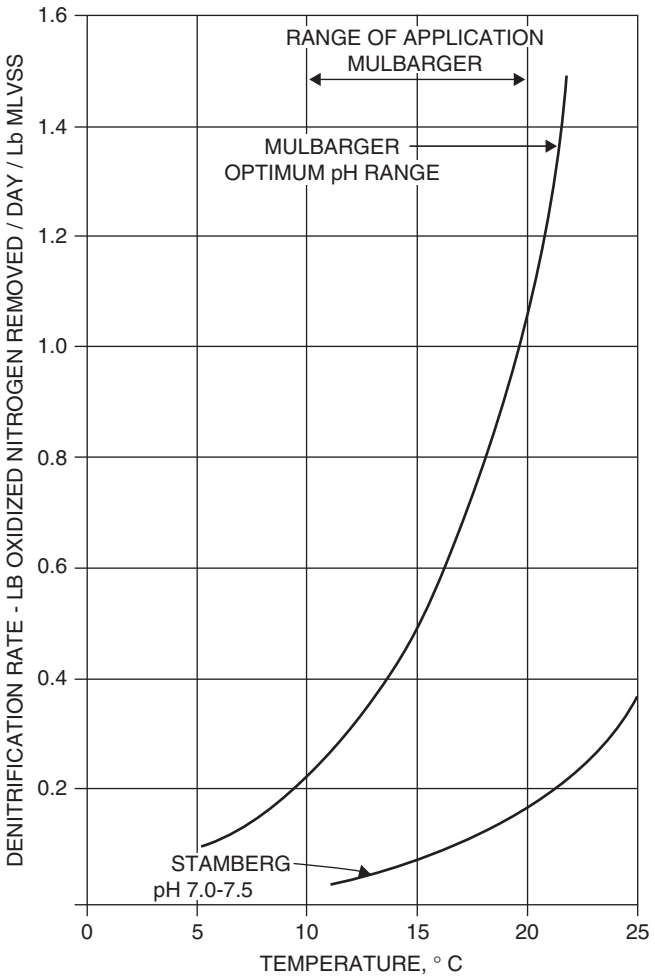

Fig. 1.20. Effect of temperature on rate of denitrification (53).

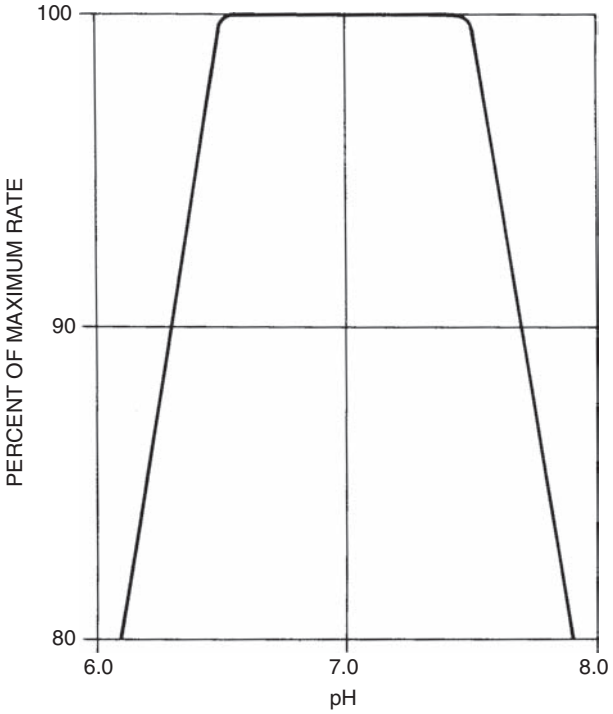

Fig. 1.21. Percent of maximum rate of denitrification versus pH (53).

shows the kinetics of the denitrification reaction in relation to temperature for a given pH range.

6.1. Effect of pH

Studies have indicated that optimum pH for the denitrifying organisms is in the range of 6.5 to 7.5, the same as for most saprophytic bacteria. Figure 1.21 shows the corrections that must be applied to the permissible tank loading when the pH is different from the optimum range.

That the pH of the effluent from the nitrifying units may exceed 7.5 at some time during a year is no particular problem, because carbon dioxide generated from oxidation of carbonaceous matter in the denitrification unit quickly reduces the pH into the favorable range below 7.5. There is no need for addition of chemicals to control pH.

6.2. MLSS and MLVSS

Experience has shown that denitrifying biosolids have settling properties comparable to biosolids from activated sludge. It seems reasonable to assume, therefore, that mixed-liquor solids in the range of 2,000 to 3,000 mg/L can be maintained without excessive rates of returning sludge. The volatile matter in denitrifying biosolids is about two-thirds of its total suspended solids.

6.3. Effect of Temperature

Reference to Fig. 1.20 will show that the minimum temperature to be allowed for will play a great role in determining the size of the denitrification tanks, as well as the MLVSS that can be carried in the system.

Systems that are usually designed for operation at a minimum temperature of 10°C would have more than twice the tankage needed at 20°C. For this reason, good design will allow for idle operation of part of the capacity during the warm months of the year. A design similar to that shown for the nitrification system in Fig. 1.11 is recommended. The tankage allowance must be considerably more generous—possibly three or four times as great if complete denitrification is to be required in the winter months. However, it is questionable whether denitrification will be needed during the low-temperature months of the year, because of the flushing action of high river flows during the spring months.

6.4. Size of Denitrification Tank

The denitrification tank layout should assure that the plug-flow mixing model is followed as closely as possible, because nitrates are not adsorbed by biological growths and detention periods may be quite short. Whether covered tanks are required to minimize absorption of oxygen from the atmosphere is a matter of conjecture. There is evidence to indicate that properly designed denitrification units can be made to seal themselves by formation of a floating scum. In any event, airtight or walk-in covers are to be avoided, because nitrogen and carbon dioxide are both released during the denitrification reaction. Figures 1.21 and 1.22 may be used to compute the size of the denitrification tanks.

6.5. Carbonaceous Matter

Effluents from nitrifying units are exceptionally free of carbonaceous BOD. For this reason, denitrification is very slow unless a readily oxidizable source of carbonaceous matter is added. Methyl alcohol (methanol) is the cheapest commercial source of carbonaceous matter. Glucose (corn sugar) is the next cheapest source. Methanol is preferable because it is more completely oxidized than glucose and, consequently, produces less sludge for disposal.

In some areas, nitrogen-deficient industrial wastes, such as brewery wastes, might be available and suitable for use. All such waste materials should be used as well as the bypassing of a small stream of original wastewater directly to the denitrification tank before considering the addition of methanol.

When methanol is used for denitrification the basic reaction involved is

$$5CH_3OH + 6H^+ + 6NO_3^- \rightarrow 5CO_2 + 3N_2 + 13H_2O \tag{72}$$

$$(5 \times 32) = 160 \qquad (6 \times 14) = 84$$

From the foregoing equation and weight relationships, it might be concluded that each pound of nitrate nitrogen would require about 2 lb of methanol for its reduction, which is true, but some of the methanol is used to produce new cell growth (biosolids) as follows:

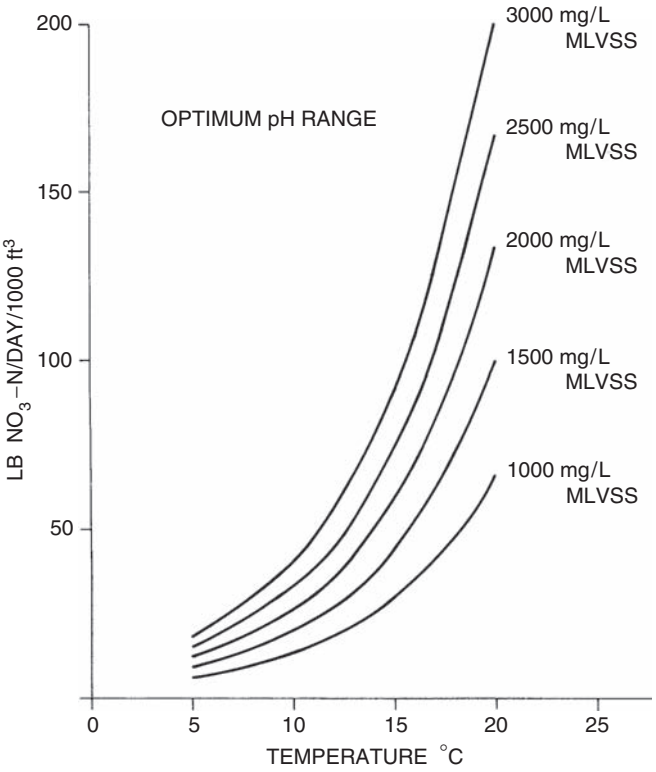

Fig. 1.22. Permissible denitrification tank loadings (53).

$$(CH_3OH)_x \rightarrow CO_2 + (CH_2O)_x + H_2O \tag{73}$$

Also, nitrified effluents normally carry some dissolved oxygen (DO) into the denitrification tank and some DO may enter the mixture as a result of agitation. This increases the amount of methanol required. An equation commonly used to estimate methanol requirements is:

$$\text{lb/d methanol} = 2.47\,\text{lb } NO_3\text{–N} + 1.53\,\text{lb } NO_2\text{–N} + 0.87\,\text{lb DO} \tag{74}$$

Reports indicate that from 3 to 4 lb of methanol/lb of nitrate nitrogen are required to consume DO and leave sufficient amount to reduce the nitrate to nitrogen gas.

The amount of methanol fed must be very closely controlled by a system such as that shown in Fig. 1.23 to insure that enough is fed to reduce the nitrates and to avoid an excess. Any excess is not only a waste of chemical; it creates an undesirable residual BOD.

6.6. Other Requirements

Equipment: The contents of the denitrification tanks are mixed with underwater mixers comparable to those used in flocculation tanks. The energy provided must be enough to keep

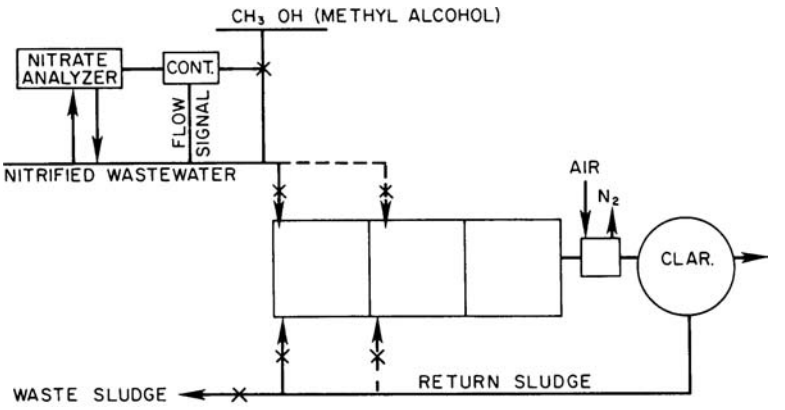

Fig. 1.23. Methyl alcohol feeding system for denitrification tanks (53).

the MLSS in suspension, but must be controlled to prevent pickup of atmospheric oxygen as much as possible, unless the tanks are covered or some other method is used to exclude contact with the air.

Power: Power requirements of 1/4 to 1/2 hp/1,000 ft^3 have been found to be adequate.

Nitrogen release: The denitrification reaction results in the formation of carbon dioxide and nitrogen gas. Both have limited solubility in water, especially the latter. Because of the gentle mixing used in the denitrification tanks, the mixed liquor leaving the tanks is supersaturated with nitrogen, and possibly carbon dioxide. As a result, gas bubbles tend to form and adhere to the MLSS and inhibit settling in the final clarifier. Supersaturated conditions can be relieved by using an aeration tank or aerated open tanks (77). It is recommended that from 5 to 10 min detention be provided at peak flow. Such a facility will also provide the ability to remove small amounts of excess methanol. Another alternative is to take advantage of the supersaturated conditions by using flotation (78, 79) rather than sedimentation for separation of biosolids from the denitrification tank effluent.

Settling tanks: Experience indicates that the settling properties of denitrifying biosolids, following relief of supersaturation, are very similar to conventional activated sludge. Tank depths of 12 to 15 ft are recommended, and surface overflow rates should not exceed 1,200 gal/ft^2/d at peak flows. MLSS concentrations greater than 2,500 mg/L may require larger tanks owing to the higher settling-tank solids loadings. A suction-type sludge collector is recommended for large circular tanks. Long rectangular tanks should be equipped with mid tank sludge-drawoff systems. Skimming facilities should be provided on the settling tanks and provisions should be made for returning the scum to the denitrification tank when desired.

Biosolids: Capability of returning biosolids to the denitrification tank of up to at least 50% and preferably of up to 100% of average flow is recommended. Provision should be made for periodic wasting of biosolids from the denitrification systems similar to that used for carbonaceous systems. Normally, the biosolids should be wasted to mix with primary and/or waste-activated sludge and be disposed of with them. The waste-biosolids line, however, should be designed to transport biosolids to the nitrification tank when desired. It is reported

Table 1.6
Expected effluent quality at 10°C (53)

Component	Concentration (mg/L)
Suspended solids	10
BOD	5
Organic-N	1.0
NH_3-N	0.5
NO_3-N	0.5
Total N	2.0

that about 0.2 lb of biosolids will be generated for each pound of methanol fed. This would correspond to about 0.7 lb/lb of nitrate nitrogen reduced.

Effluent quality: Table 1.6 indicates the expected effluent quality from a nitrification–denitrification system designed for operation at 10°C wastewater temperatures. At warmer temperatures improved quality can be expected.

Thus, it seems that more than 90% removals of total nitrogen can be achieved in actual practice.

7. DESIGN EXAMPLES

7.1. Example 1

A biochemical reaction that follows first-order kinetics has a measured reaction rate constant of 15/d at 20°C.

1. Using $\alpha = 1.072$, calculate the rate constant, k_1 at 25°C
2. Find the corresponding required time for the substrate to decrease in concentration from 200 to 20 mg/L

Solution:

1. The rate constant, k_1 at 25°C

$$r_T = r_{20}\alpha^{(T-20)} \tag{2}$$

$$k_{25} = 15(1.072)^{(25-20)}$$

$$k_{25} = 15 \times 1.416 = 21.2/\text{d}$$

2. The required time, t

$$S = S_o e^{-k_1 t} \tag{17}$$

$$\ln \frac{S}{S_o} = -k_1 t$$

$$t = -\frac{1}{k_1} \ln \frac{S}{S_o}$$

$$t = -\frac{1}{21.2} \ln \frac{20}{200}$$

$$t = 0.11 \text{ day}$$

$$t = 0.11 \times 24 = 2.64 \text{ h}$$

7.2. Example 2

An industrial wastewater stream, which is produced at a rate of 476 m^3/d, contains a pollutant concentration of 300 mg/L. A bench-scale study showed that the pollutant removal follows a first-order kinetic reaction, where $k_1 = 2.5$/d.

If the National Standards require a minimum removal of 85% of the pollutant before the effluent is allowed to be discharged into the sewerage system, determine the required size of a completely mixed reactor that can accomplish the job.

Solution:

$$\text{Change} = \text{input} - \text{output} - \text{sink}$$

$$v\frac{dS}{dt} = QS_o - QS + Vr$$

For a first-order reaction, $r = -k_1 S$

$$v\frac{dS}{dt} = QS_o - QS + V(-k_1 S)$$

At steady state, $\frac{dS}{dt} = 0$

$$QS_o - QS - Vk_1 S = 0$$

Because $V = Qt$

$$QS_o - QS - Qtk_1 S = 0$$

$$S_o - S - k_1 St = 0$$

$$t = \frac{S_o - S}{k_1 S}$$

$$t = \frac{1}{k_1}\left[\frac{S_o}{S} - 1\right]$$

$$t = \frac{1}{2.5}\left[\frac{300}{300(1.00 - 0.85)} - 1\right]$$

$$t = \frac{5.67}{2.5} = 2.27 \text{ day}$$

$$V = Qt$$

$$V = 476\,\text{m}^3/\text{d} \times 2.27\ \text{day}$$

$$V = 1{,}080\,\text{m}^3$$

7.3. Example 3

It is required to design a nitrification tank using US EPA criteria under the following operating conditions:

(a) Design flow = 10 MGD
(b) Average BOD_5 = 30 mg/L
(c) Average concentration of NH-N = 15 mg/L
(d) Minimum temperature 10°C
(e) Operating pH = 7.8
(f) MLVSS concentration = 1,500 mg/L

Determine:

1. Required tank size
2. Required detention time
3. Oxygen requirements

Solution:

Average NH_3 load = $10 \times 8.34 \times 15 = 1{,}250$ lb/d
Maximum NH_3 load = $1250 \times 1.5 = 1{,}870$ lb/d
BOD_5 load = $10 \times 8.34 \times 30 = 2{,}500$ lb/d

1. Tank Size
From Fig. 1.13 (at 10°C, MLVSS = 1500 mg/L and optimum pH 8.4) volumetric loading = 8.2 lb/d/1,000 ft^3

$$\text{Tank volume} = \frac{1870}{8.2} \times 1000 = 228{,}000\,\text{ft}^3$$

From Fig. 1.14 (at pH 7.8) volume adjustment = 0.88

$$\text{Tank Size} = \frac{228{,}000}{0.88} = 260{,}000\,\text{ft}^3$$

2. Detention time

$$\text{Detention time} = \frac{(260,000)(24)(7.48)}{(10)(10^6)} = 4.67\,\text{h}$$

3. Oxygen requirements
Each lb of ammonia-nitrogen that is nitrified requires 4.6 lb of oxygen
Oxygen requirement for NH_3 oxidation = $1{,}870 \times 4.6 = 8{,}650$ lb/d
$BOD_5 = 2/3\ BOD_{Ultimate}$
Oxygen requirement to satisfy remaining BOD = $2500 \times 1.5 = 3750$ lb/d
Total oxygen requirement = $8{,}650 + 3{,}750 = 12{,}400$ lb/d

7.4. Example 4

It is required to design a denitrification tank using US EPA criteria under the following operating conditions:

1. Design flow = 10 MGD
2. Average NO_3-N + NO_2-N concentration = 15 mg/L
3. Minimum temperature 10°C
4. Expected operating pH = 7.3
5. MLVSS = 2,000 mg/L
6. Assume complete conversion is desired

Determine:

1. Required tank size
2. Required detention time

Solution:

Average NO_3-N + NO_2-N loading = $10 \times 8.35 \times 15 = 1{,}250$ lb/d
Peak NO_3-N + NO_2-N loading = $1{,}250 \times 1.5 = 1{,}870$ lb/d

1. Required tank size
   From Fig. 1.22, tank loading {at 10°C, MLVSS = 2,000 and optimum pH}= 27 lb/1,000 ft^3

$$\text{Tank volume } \frac{1870}{27}\,1000 = 70{,}000\ \text{ft}^3$$

   From Fig. 1.21, (at pH 7.3) volume adjustment = 1.0

$$\text{Tank volume} = 70{,}000 \times 1.0 = 70{,}000\ \text{ft}^3$$

2. Required detention time

$$\text{Detention period} = \frac{(70{,}000)(7.48)(24)}{(10)(10^6)} = 1.26\ \text{h}$$

NOMENCLATURE

b = temperature coefficient ($e^b = \alpha$)
c = a dimensionless number, $c = K_R/K_T$
dS/dt = rate of substrate (NH_3-N) utilization, mg/L/d
E = total enzyme concentration, mg/L
[ES] = concentration of the enzyme-substrate complex, mg/L
F/M = Food to microorganisms' ratio
$k = (q_{max})$ = maximum rate constant of NH_3-N utilization per unit weight of microorganisms, mg/L NH3-N/mg/L MLVSS/d or 1/d
$k_o = Xq_{max}$ = zero-order rate constant, mg/L/d
$k_1 = Xq_{max}/K_s$ = first-order biodegradation rate constant, 1/d
k_{20} = maximum rate constant at 20°C, 1/d
k_T = maximum rate constant at temperature T, 1/d
K_d = constant decay coefficient

K_i = inhibition coefficient
K_R = the microscopic dissociation constants of a substrate (S) bound to a site in the R state, mg/L
K_s = half-velocity coefficient or half-saturation constant, mg/L
K_T = the microscopic dissociation constants of a substrate (S) bound to a site in the T states, mg/L
L = allosteric constant
= equilibrium constant for the $R_0 \leftrightarrow T_0$ transition
n = number of binding sites or protomers
= an interaction coefficient
q = specific substrate utilization rate, mg/L NH3-N/mg/L MLVSS/d or 1/d
q_{max} = maximum specific substrate utilization rate = μ_{max}/Y, mg/L NH3-N/mg/L MLVSS/d or 1/d
Q = flow rate, MGD, gal/h, gal/min, ft^3/s
Q_o = initial or influent flow rate, MGD, gal/h, gal/min, ft^3/s
Q_e = effluent flow rate, MGD, gal/h, gal/min, ft^3/s
Q_r = return line flow rate, MGD, gal/h, gal/min, ft^3/s
Q_w = wasted biosolids flow rate, MGD, gal/h, gal/min, ft^3/s
r_s = rate of soluble substrate utilization, 1/d
r_T = biodegradation rate at temperature T, 1/d
r_{20} = biodegradation rate at 20°C, 1/d
R = recycling ratio
S = substrate concentration at any time t, mg/L
S_o = initial or influent substrate concentration at $t = 0$, mg/L
S_e = effluent substrate concentration, mg/L
S_{min} = minimum substrate concentration, mg/L
S_r = substrate concentration in returned biosolids, mg/L
S_w = substrate concentration in wasted biosolids, mg/L
SF = safety factor
SVI = sludge volume index
t = time, h or day
T = temperature, °C
T, R = represent the two states or conformations of the enzyme
$v = qX$ = rate of substrate utilization, mg/L NH_3-N/d or mg/L/d
$V_m = kX = (q_{max}X)$ maximum rate of ammonia utilization, mg/L NH_3-N/d or mg/L/d
V = Volume, ft^3 or gal
X = concentration of mixed liquor volatile suspended solids (MLVSS), mg/L
= biomass concentration, mg/L
X_e = biomass concentration in effluent, mg/L
X_r = biomass concentration in the returned biosolids, mg/L
X_w = wasted biosolids concentration, mg/L
Y = the fraction of sites bound with substrate
= saturation function, a dimensionless number

Y = growth yield
Y_{max} is the maximum growth yield
Y_{obs} = observed growth yield
Z = the reduced substrate concentration, a dimensionless number = S/K_R
α = temperature-activity coefficient
μ = specific growth rate
μ_{max} = maximum specific growth rate
θ = Mean hydraulic retention time for the aeration tank, h
θ_x = Mean cell retention time or solids retention time (SRT), day
$(\theta_x)_{min}$ = minimum value of solids retention time (SRT), day

REFERENCES

1. G. Bitton, *Wastewater Microbiology*, Wiley-Liss, New York (1999).
2. L. D. Benefield and C. W. Randall, *Biological Process Design for Wastewater Treatment*, Prentice-Hall, New Jersey (1980).
3. B. E. Rittmann and P. L. McCarty, *Environmental Biotechnology*, McGraw-Hill, New York (2001).
4. M. D. LaGrega, P. L. Buckingham, and J. C. Evan, *Hazardous Waste Management*, McGraw-Hill, New York (2001).
5. W. W. Eckenfelder, *Industrial Water Pollution Control*, McGraw-Hill, New York (2000).
6. R. L. Droste, *Theory and Practice of Water and Wastewater Treatment*, Wiley, New York (1997).
7. Metcalf & Eddy, *Wastewater Engineering*, McGraw-Hill, New York (2003).
8. S. J. Pirt, The maintenance energy of bacteria in growing cultures, *Proceedings of the Royal Society of London*, B163, 224–231 (1965).
9. J. Chang, P. Chudoba, and B. Capdeville, Determination of the maintenance requirement of activated sludge, *Water Science and Technology*, 28, 139–142 (1993).
10. Y. Liu, and J. H. Tay, Interaction between catabolism and anabolism in the oxidative assimilation of dissolved organic carbon, *Biotechnology Letters*, 22, 1521–1525 (2000).
11. D. Herbert, R. Elsworth, and R. C. Telling, The continuous culture of bacteria: a theoretical and experimental study, *Journal of General Microbiology*, 114, 601 (1956).
12. A. W. Lawrence and P. L. McCarty, Unified basis for biological treatment design and operation, *Journal of Sanitary Engineering*, 96, 757 (1970).
13. APHA, *Standard Methods for the Examination of Water and Wastewater*, 20th ed., American Public Health Association, Washington, DC (1998).
14. W. W. Eckenfelder and Y. Argaman, principles of biological and physical/chemical nitrogen removal. In: *Phosphorus and Nitrogen removal from Municipal Wastewater*, R. I. Sedlak (ed.), pp. 3–42, Lewis Publishers, New York (1991).
15. L. K. Wang, D. Veilkind, and M. H. Wang, Computer-aided mathematical modeling of stream purification capacity, Part I: Nonlinear DO model, *Institute of Environmental Sciences 1976 Proceedings*, 553 (1976).
16. L. K. Wang, C. P. C. Poon, and M. H. Wang, Control tests and kinetics of activated sludge process, *Water, Air and Soil Pollution*, 8, 315–351 (1977).

17. S. F. Yang, J. H. Tay, and Y. Liu, A novel granular sludge sequencing-batch reactor for organic and nitrogen removal from wastewater, *Journal of Biotechnology*, 106, 77–86 (2003).
18. P. Chudoba, Etude et Intérêt du Découplage Energetique dans les Processus D'épuration des Eaux par Voie Biologique Procédé OSA, Ph.D. thesis, Institut National des Sciences Appliquées, Toulouse, France (1991).
19. L. K. Wang, M. H. Wang, and D. B. Dahm, Design, Cost Estimation and Optimization of Sewage Collection and Treatment Systems for Housing Development in the Glenwood, New York Area, Technical Report No. ND-5390-M-1, Calspan Corporation, Buffalo, New York (1974).
20. J. R. McWhirter, Oxygen and activated sludge process, In: *The Use of High-Purity Oxygen in the Activated Sludge Process*, Vol. 1, J. R. Mcwhirter (ed.), pp. 25–62, CRC Press, Boca Raton, FL (1978).
21. H. Roques, B. Capdeville, J. C. Seropian, and H. Grigoropoulou, Oxygenation by hydrogen peroxide of the fixed biomass used in biological water treatment, *Water Research*, 18, 103–110 (1984).
22. B. Abbassi, S. Dullstein, and N. Rabiger, Minimization of excess sludge production by increase of oxygen concentration in activated sludge flocs: experimental and theoretical approach, *Water Research*, 34, 139–146 (2000).
23. A. G. Boon, and D. R. Burgess, Treatment of crude sewage in two high-rate activated sludge plants operated in series, *Water Pollution Control*, 74, 382 (1974).
24. T. R. Stall, and J. H. Sherrard, Effect of wastewater composition and cell residence time on phosphorus removal in activated sludge, *Journal of Water Pollution Control Federation*, 48, 307–322 (1976).
25. R. Wunderlich, J. Barry, D. Greenwood, and C. Carry, Start-up of a high-purity, oxygen-activated sludge system at the Los Angeles County SanitationDistricts' Joint Water Pollution Control Plant, *Journal of the Water Pollution Control Federation*, 57, 1012–1018 (1985).
26. N. J. Horan, Biological Wastewater Treatment Systems, Wiley, Chichester (1990).
27. Y. Liu, and J. H. Tay, Strategy for minimization of excess sludge production from the activated sludge process, *Biotechnology Advances*, 19, 97–107 (2001).
28. Y. Liu, Chemically reduced excess sludge production in the activated sludge process, *Chemosphere* 50, 1–7 (2003).
29. N. K. Shammas, *Optimization of Biological Nitrification*, Ph.D. dissertation, Microfilm Publication, University of Michigan, Ann Arbor, MI (1971).
30. K. M. Mackenthun, A Review of algae, lake weeds, and nutrients, *Journal of the Water Pollution Control Federation*, 34, 1077 (1962).
31. K. Wuhrmann, Objectives, technology, and of results of nitrogen and phosphorus removal processes, *University of Texas, Water Resources Symposium* No. 1, 21 (1968).
32. W. W. Eckenfelder, Jr., A design procedure for biological nitrification and denitrification, *Chemical Engineering Progress Symposium Series*, 63(78), 230 (1967).
33. P. M. Sutton, K. L. Murphy, and B. E. Jank, Kinetic studies of single sludge nitrogen removal systems, *Progress in Water Technology*, 10, 241 (1978).
34. D. F. Bishop, J. A. Heidman, and J. B. Stamberg, Single-stage nitrification-denitrification, *Journal of the Water Pollution Control Federation*, 48, 520 (1976).
35. P. M. Sutton, K. L. Murphy, and B. E. Jank, Design considerations for integrated nutrient removal systems, *Progress in Water Technology*, 10, 469 (1978).
36. P. A. Vesilind (ed.), *Wastewater Treatment Plant Design*, Water Environment Federation, Alexandria, VA (2003).

37. P. H. Jones and H. N. Sabra, Effect of systems solids retention time on nitrogen removal from activated sludge, *Water Pollution Control*, 79, 106 (1980).
38. C. S. Huang and N. E. Hopson, Nitrification rate in biological processes. *Journal of Environmental Engineering Division, American Society of Civil Engineers* EE2, 100, 409 (1974).
39. P. M. Sutton, T. R. Bridle, W. K. Bedford, and J. Arnold, Nitrification and denitrification of an industrial wastewater, *Journal of the Water Pollution Control Federation*, 53, 176 (1981).
40. C. W. Randall and D. Buth, Nitrite build-up in activated sludge resulting from temperature effects, *Journal of the Water Pollution Control Federation*, 56, 1039 (1984).
41. N. K. Shammas, Biocontactors for wastewater reuse, kinetic approach for achieving the required effluent quality, *First Saudi Engineering Conference*, Jeddah, KSA, May (1983).
42. A. P. Sincero and G. A. Sincero, *Environmental Engineering – A Design Approach*, Prentice-Hall, Upper Saddle River, NJ (1996).
43. M. H. Gerardi, *Nitrification and Denitrification in the Activated Sludge Process*, Wiley Inter-Science, New York, December (2001).
44. N. K. Shammas, Interactions of temperature, pH and biomass on the nitrification process, *Journal of the Water Pollution Control Federation*, 58, 1, 52–59, January (1986).
45. A. L. Downing et al., Nitrification in the activated sludge process, *Journal of the Institute of Sewage Purification*, 130 (1964).
46. G. Knowles et al., Determination of kinetic constants for nitrifying bacteria in mixed culture with the aid of an electronic computer, *Journal of General Microbiology*, 38, 263 (1965).
47. F. E. Stratton and P. L. McCarty, Prediction of nitrification effects on the dissolved oxygen balance of streams, *Environmental Science and Technology*, 1, 405 (1967).
48. H. A. Painter and K. Jones, The use of the wide-bore dropping- mercury electrode for the Determination of rates of oxygen uptake and of oxidation of ammonia by microorganisms, *Journal of Applied Bacteriology*, 26, 471 (1963).
49. H. E. Wild, C. N. Sawyer, and T. C. McMahon, Factors affecting nitrification kinetics, *43rd Annual Conference, Water Pollution Control Federation*, Boston, MA, October 9 (1970).
50. D. F. Bishop et al., Single-stage nitrification-denitrification, *Journal of the Water Pollution Control Federation*, 48, 520 (1976).
51. A. L. Downing, Factors to be considered in the design of activated sludge plants, *Proceedings of Water Resources Symposium No. 1*, University of Texas Press, Austin, TX, 190 (1968).
52. A. M. Buswell et al., Laboratory studies on the kinetics of the growth of nitrosomonas with relation to the nitrification phase of the BOD test, *Applied Microbiology*, 2, 21 (1954).
53. US EPA, *Nitrification and Denitrification Facilities-Wastewater Treatment*, Environmental Protection Agency, Technology Transfer, EPA-625/4-73-004a, August (1973).
54. N. K. Shammas, An allosteric kinetic model for the nitrification process, *Proceedings of Tenth Annual Conference of Water Supply Improvement Association*, Honolulu, HI, pp. 1–30, July (1982).
55. J. Monod, J. Wyman, and J. P. Changeux, On the nature of allosteric transitions: a plausible model, *Journal of Molecular Biology*, 12, 88 (1965).
56. C. Frieden, Treatment of enzyme kinetic data, II the multisite case: comparison of allosteric models and a possible new mechanism, *Journal of Biological Chemistry*, 242, 4045 (1967).
57. M. V. Volkenstein and B. N. Goldstein, Allosteric enzyme models and their analysis by the theory of graphs, *Biochemica et Biophysica Acta*, 115, 478 (1966).
58. B. R. Rabin, Co-operative effects in enzyme catalysis: a possible kinetic model based on substrate-induced conformation isomerization, *Biochemical Journal*, 102, 22c (1967).

59. J. R. Sweeny and J. R. Fisher, An alternative to allosterism and cooperativity in the interpretation of enzyme kinetic data, *Biochemistry*, 7, 561 (1968).
60. D. Koshland, G. Nemethy and D. Filmer, Comparison of experimental binding data and theoretical models in proteins containing subunits, *Biochemistry*, 5, 365 (1966).
61. E. Kvamme and A. Pihl (eds.), Regulation of enzyme activity and allosteric interactions, *Proceedings of the 4th Meeting of the Federation of European Biochemical Societies*, Oslo (1967), Academic Press, New York (1968).
62. T. E. Barman, *Enzyme Handbook*, Springer-Verlag, New York (1969).
63. H. Boring and A. Horon, Analysis of kinetic data of allosteric enzymes by a linear plot, *FEBS Letters*, 3, 325 (1969).
64. B. Formby and J. Clausen, Allosteric and kinetic properties of a K-dependent acylphosphatase in rat brain synaptosomes and its possible relation to $(Na^+ + K^+)$ ATPase, *Hoppe-Seyler's Zeitschrift Für Physiologische Chemie*, 350, 973 (1969).
65. K. Kirschner, Temperature-jump relaxation with an allosteric enzyme: glyceraldehyde-3-phosphate dehydrogenase, *Proceedings of the 4th Meeting of FEBS*, Academic Press, New York, 39 (1968).
66. P.L. Ipata and G. Cercignani, The effect of pH on the allosteric properties of sheep brain. *FEBS Letters*, 7, 129 (1970).
67. WEF, Technology Assessments: Nitrogen Removal Using Oxidation Ditches, Water Environment Federation, July (2000).
68. R. I. Sedlak, *Phosphorus and Nitrogen Removal from Municipal Wastewater: Principles and Practice*, 2nd ed., Lewis Publishers, New York, October (1991).
69. B. R. Blicker, *Process Optimization of an On-Site Wastewater Treatment System for Nitrogen Removal*, M.S. thesis, Environmental Engineering, Montana State University, December (1997).
70. M. Debabrata, Hybrid reactor system for wastewater treatment—application and approach of modeling, *International Journal of Environment and Pollution (India)*, 21, 2, 105–131 (2004).
71. L. K Wang, N. C Pereira, and Y. T. Hung (eds.), *Biological Treatment Processes*, Humana Press, Totowa, NJ (2005).
72. Y. Liu, J. H. Tay, and Y. T. Hung, Biological nitrification and denitrification. In: *Biological Treatment Processes*, L. K Wang, N. C Pereira, and Y. T. Hung (eds.), Humana Press, Totowa, NJ (2005).
73. B. Halling-Soresen and S. E. Jorgensen, *Removal of Nitrogen Compounds from Wastewater*, Elsevier, Amsterdam, (1993).
74. H. Constantin, S. Raoult, W. Montigny, and M. Fick, *Environmental Technology*, 17, 831–840 (1996).
75. EvTEC, *Wastewater Denitrification Technologies*, Pennsylvania Department of Environmental Protection, http://www.cerf.org/evtec/eval/padep.htm, April (2001).
76. INI, *Workshop on Advanced Approaches to Quantify Denitrification*, International Nitrogen Initiative, Sponsored by EPA, Workshop Final Report May (2004).
77. L. K Wang, Y. T. Hung, and N. K. Shammas (eds.), *Advanced Physicochemical Treatment Processes*, Humana Press, Totowa, NJ (2005).
78. L. K. Wang, E. Fahey, and Z. Wu, Dissolved air flotation. In *Physicochemical Treatment Processes*, L. K Wang, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ (2005).
79. L. K Wang, Y. T. Hung, and N. K. Shammas (eds.), *Physicochemical Treatment Processes*, Humana Press, Totowa, NJ (2005).

2
Vertical Shaft Bioreactors

Nazih K. Shammas, Lawrence K. Wang, Jeffrey Guild, and David Pollock

CONTENTS

Abstract The Vertical Shaft Bioreactor (VSB) treatment system is essentially a high rate activated sludge process capable of operating at food to microorganism ratios (F/M) of between 0.5 and 2.0 kg BOD_5/kg MLVSS/d. These extremely high loadings are achievable because of the capability of the system to carry and maintain mixed liquor volatile suspended solids (MLVSS) concentration values between 5.000 and 10,000 mg/L. As a result, a much lower volume (aeration period) is required than in the conventional activated sludge process. The process consists of a vertical subsurface reactor shaft 0.75 to 6 m in diameter and 75 to 125 m deep, with hydraulic mean residence times in the order of 60 min.

The following aspects of the VSB process are covered: Process description, technical development, vertical bioreactor system and its variations, process theory and design basis, process design, operation, and maintenance, comparison with equivalent technologies, and case studies.

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_2

Key Words Biological treatment • flotation and design • industrial treatment • municipal treatment • vertical shaft bioreactor (VSB).

1. PROCESS DESCRIPTION

The vertical shaft bioreactor (VSB) treatment system is essentially a high-rate activated sludge process capable of operating at food to microorganism ratios (F/M) of between 0.5 and 2.0 kg BOD_5/kg MLVSS/d (1). These extremely high loadings are achievable because of the capability of the system to carry and maintain mixed liquor volatile suspended solids (MLVSS) concentration values between 5,000 and 10,000 mg/L. As a result, a much lower volume (aeration period) is required than in the conventional activated sludge process (2, 3).

The process consists of a vertical subsurface reactor shaft, which in early configurations was between 90 and 150 m (300 to 500 ft) deep, with hydraulic mean residence times in the order of 60 minutes (4). The reactor is typically installed using conventional drilling equipment as illustrated in Figs. 2.1 and 2.2. In general, carbon steel shafts are used for the exterior casing (Fig. 2.3). The shafts are typically grouted with sulfate-resistant cement to allow isolation from the surrounding geological formation.

Fig. 2.1. Drilling rig and cutting bit (Courtesy NORAM Engineering and Construction Ltd.)

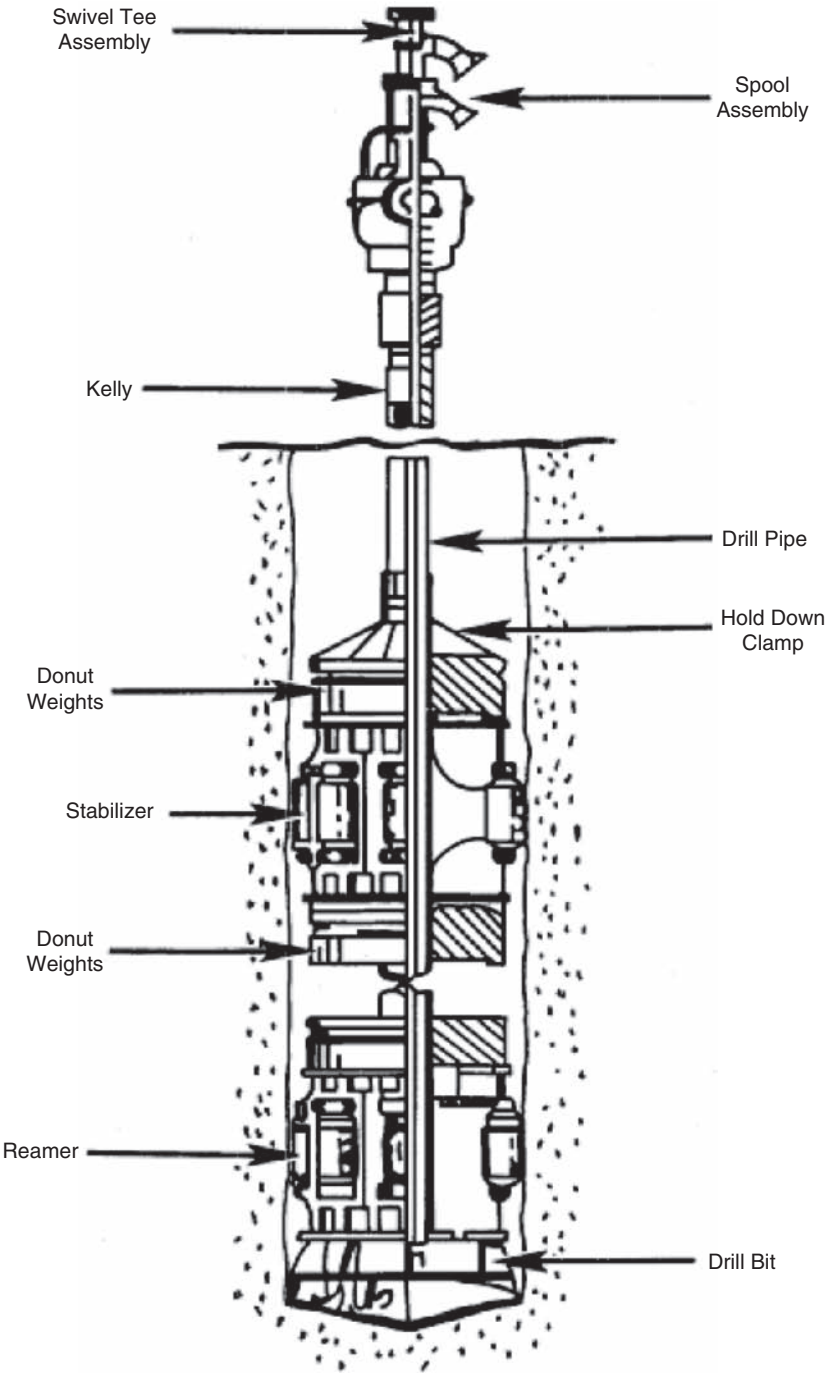

Fig. 2.2. Vertical hole drilling Assembly (4)

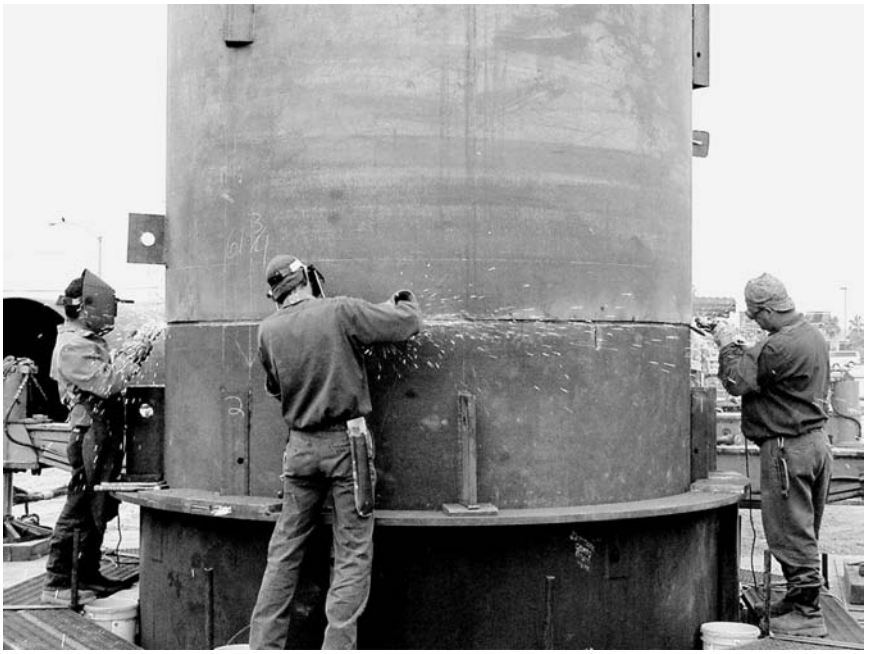

Fig. 2.3. Reactor placement (Courtesy NORAM Engineering and Construction Ltd.)

The reactor is divided basically into two sections, namely, a downflow section called a downcomer and an upflow section called a riser. In the initial reactor configuration, the raw wastewater and return sludge are introduced into the downcomer section of the reactor, and the mixed liquor is withdrawn from the riser section. Compressed air is introduced into both the downcomer and the riser sections to serve as a source of oxygen, as well as the driving force for fluid transport through the shaft. The air requirements and air injection depth are determined by taking into consideration the minimum liquid circulation velocity and BOD_5 removal requirements. In general, liquid circulation velocities between 0.9 and 1.5 m/s (3 to 5 ft/s) are maintained within the VSB (4). Depending on the operating mixed liquor volatile suspended solids (MLVSS) concentration, the effluent from the reactor can be treated for solids separation using either a flotation or sedimentation process.

In the case of domestic wastewater treatment, the raw influent wastewater generally undergoes preliminary treatment for the removal of large particles (screenings) and grit. Experience with the VSB process indicates that the process can operate successfully without primary clarification. Figure 2.4 shows a conceptual flow diagram for the treatment of domestic wastewaters using an early generation VSB. Figure 2.5 shows the general concept and hydraulic flow pattern occurring within a VSB.

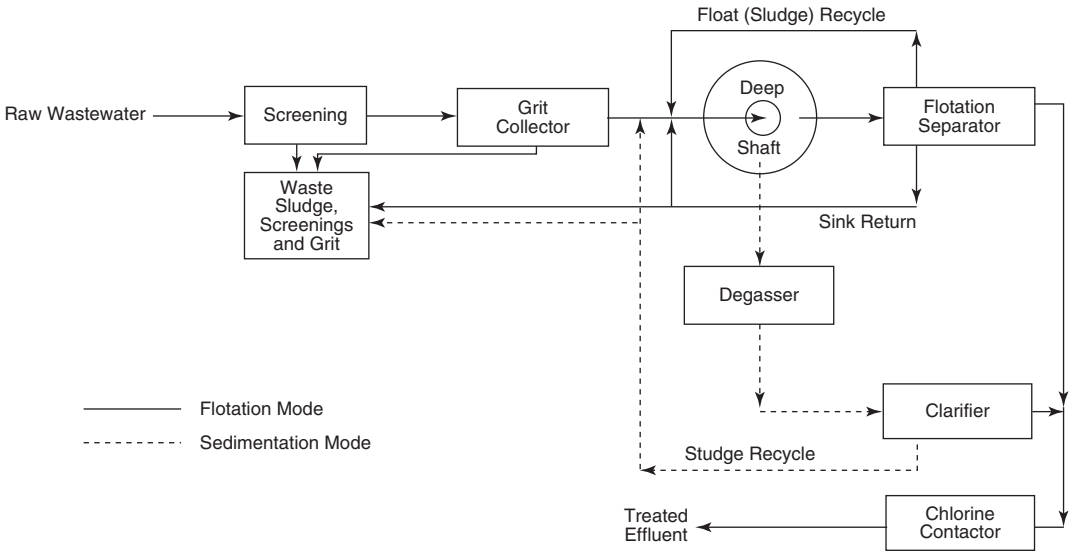

Fig. 2.4. Conceptual treatment process flow diagram (4)

2. TECHNICAL DEVELOPMENT

VSB use in biological treatment of wastewaters has its origin in the United Kingdom and was developed from research efforts for the synthesis and production of single cell protein using methanol as feedstock (5). The process required the operation of the system with high bacterial density. In order to satisfy the extremely high requirements for dissolved oxygen, Imperial Chemical Industries Limited (ICI) adopted a pressure cycle aerobic fermentor vertical shaft in which an increased hydrostatic pressure between 90 and 150 m (300 to 500 ft) was used to increase oxygen transfer capabilities. The pressure cycle fermentor used airlift principles in which the air for biochemical oxidation also provided the air for liquid circulation. An extension of this basic research and development work is the application of the process principles for wastewater treatment. Wastewater treatment application normally involves the operation of the VSB with lower bacterial density, less biodegradable substrate (BOD_5), and a slower growth rate of microorganisms than in the single-cell protein reactor. For these reasons, ICI, modified the reactor configuration and increased the typical design depth of the reactor to achieve equivalent oxygen transfer efficiency and power economy (6). In addition, ICI initiated several pilot and demonstration projects involving municipal and industrial wastewaters.

The first version of the VSB process consisted of a deep subsurface well, a head tank, and a solids separation clarifier. The unit configuration, along with the gas voidage and dissolved oxygen profiles, are presented in Fig. 2.6. Gas voidage refers to the volume fraction of entrapped gas bubbles in the mixed liquor, and can be expressed as follows:

$$\text{Gas voidage} = V_G/(V_G + V_L) \tag{1}$$

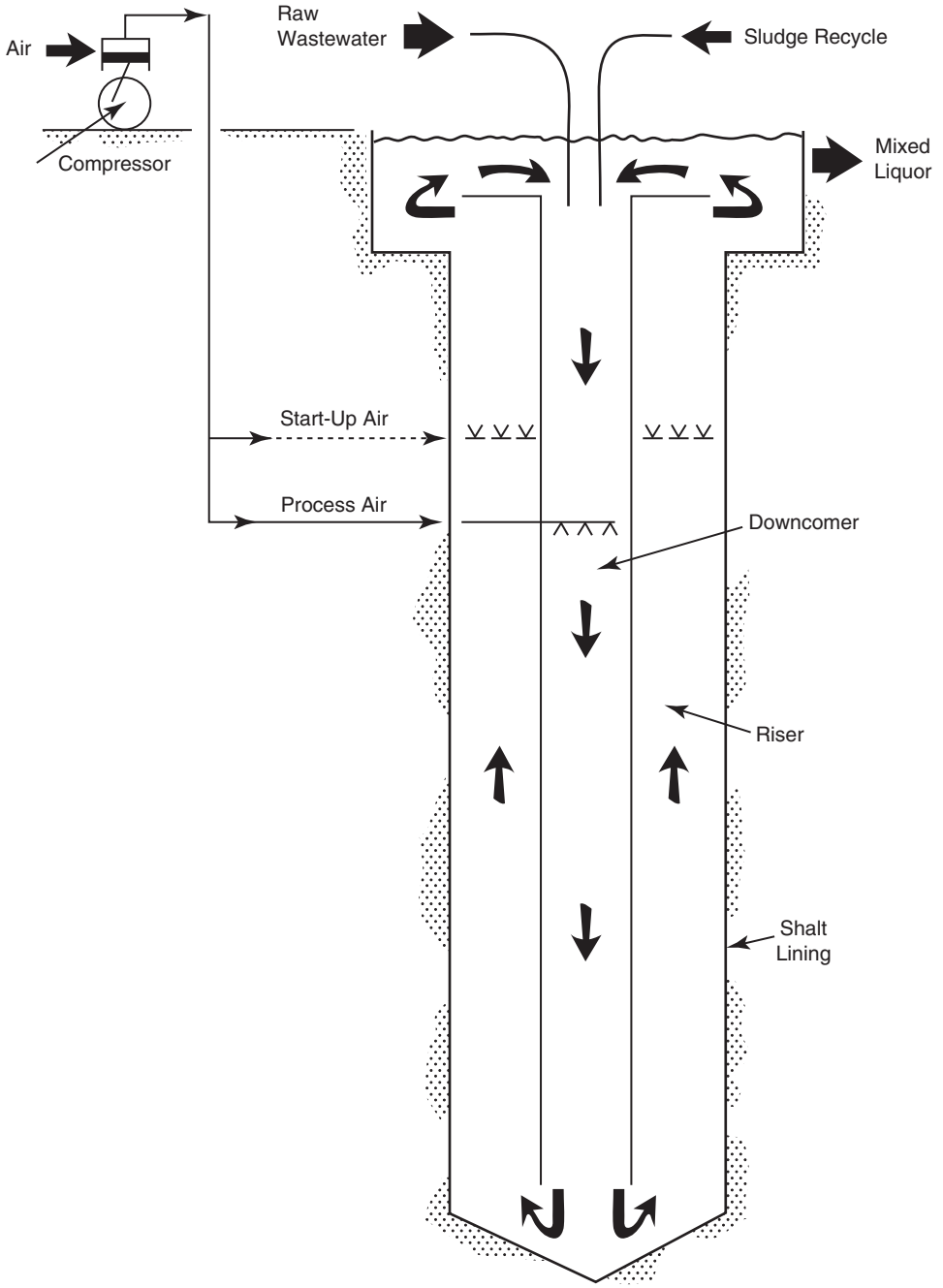

Fig. 2.5. Hydraulic flow pattern in the ICI process (4)

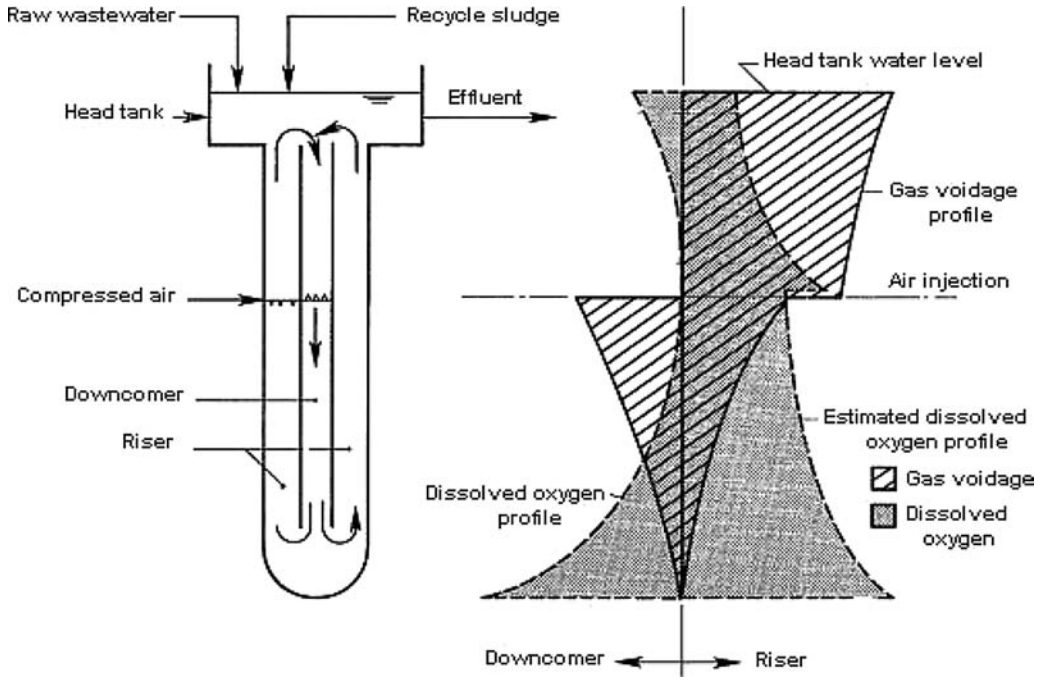

Fig. 2.6. ICI process diagram with gas voidage and dissolved oxygen profile (4)

where

V_G = Volume of gas bubbles
V_L = Volume of liquid

The gas voidage difference between the riser and the downcomer sections of the VSB is used to initiate and maintain liquid circulation within the reactor.

The ICI vertical shaft bioreactor is also divided into two concentric sections similar to the early fermentor. Raw wastewater and recycle sludge are introduced into an open head tank from which the mixed liquor flows down the downcomer and upward through the annular riser section to the head tank. Mixed liquor is also withdrawn from the head tank for solids separation and to provide for recycle sludge. Based on these operating principles, a pilot plant was started by ICI in Billingham, England during 1974. The pilot plant had a design capacity of approximately 363 m^3/d (96,000 gpd). A 39 cm (15.25 in.) diameter shaft, 130 m (426 ft) deep provided the outside shell for the VSB process (4).

During the initial operation of the pilot plant, the solids separation process consisted of a dissolved air flotation unit followed by a mechanical degasser and clarification unit. The flotation separator was included in the process to make use of the potentially available

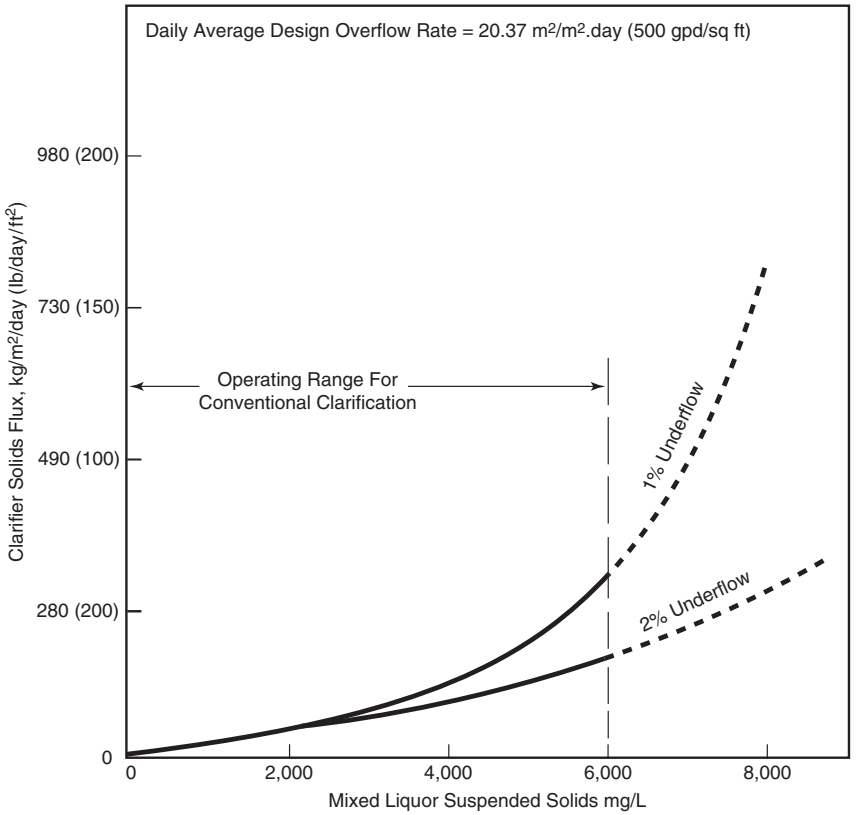

Fig. 2.7. Relationship between MLSS concentration and solids flux (4)

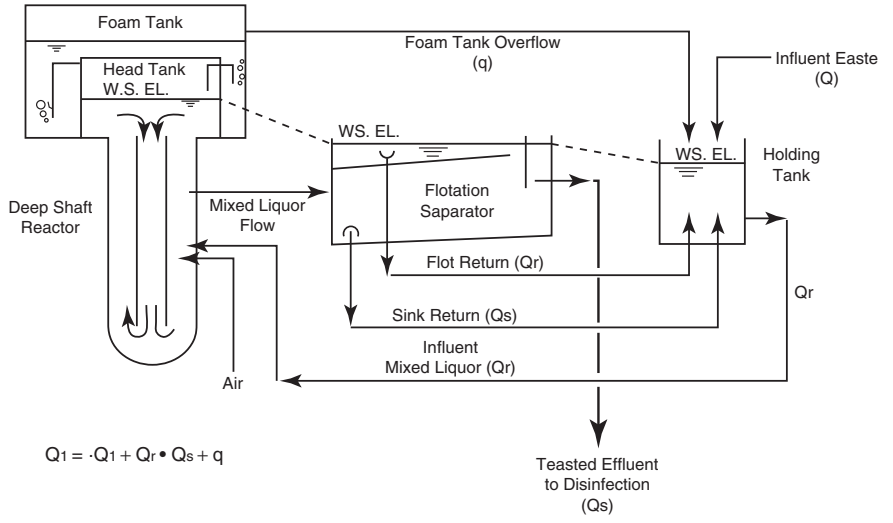

Fig. 2.8. VSB hydraulic profile (4)

dissolved gases present in the VSB mixed liquor. Subsequent experience with the VSB system indicated that the flotation unit and the mechanical degasser can be replaced with a vacuum degasser before clarification. Further testing using the clarification mode indicated that the process is capable of producing better than secondary quality effluent ($BOD_5 = 15\,mg/L$, $SS = 18\,mg/L$) when operating at mixed liquor suspended solids concentration (MLSS) values between 2,000 and 6,000 mg/L (6–8).

In summary, the process development and successful demonstration at Billingham, England created sufficient interest in Europe, North America, and Japan to warrant extensive marketing efforts. Accordingly, ICI extended process licenses to Canadian Industries Limited (CIL) of Canada for carrying out the marketing efforts in North America. Eco Technology (Eco), a division of CIL, assumed this responsibility in mid-1975 and contributed significantly to the exposure and development of this technology.

Eco recognized that the VSB volume can be significantly reduced if the overall system could be designed to operate with high mixed liquor suspended solids ($\geq$6,000 mg/L). Eco also realized that the limiting constraint for operating the system with high mixed liquor suspended solids was the gravity separation of the mixed liquor solids leaving the VSB. The use of a gravity separation process unit (e.g., clarifier) has generally been limited to MLSS concentration values below 6,000 mg/L (9, 10). This is due to the recommended design criteria for solids flux through the gravity separation unit. Figure 2.7 shows the relationship between MLSS, hydraulic overflow rate, and solids flux rate for conventional clarification.

Based on this consideration, Eco's development work was directed toward incorporating the flotation separator with the VSB process. As a result of Eco's research work, the VSB and the flotation separator unit will be operating under constant hydraulic and solids loading. The air supply requirements will also be maintained at a steady rate to achieve desired liquid circulation velocities inside the reactor. Figure 2.8 shows the typical hydraulic profile and the flow routing for the Eco-III reactor system.

When the raw wastewater flow rate exceeds the normal design conditions, the liquid level inside the head tank rises to maintain flow through the reactor. A liquid level or pressure sensor inside the head tank will actuate and increase air supply to accommodate the increased hydraulic flow or to provide additional oxygen requirements. The output signal from the liquid level control can also be interconnected with the float skimmer drive mechanism to attain increased recycle float sludge.

3. VERTREAT BIOREACTOR

VERTREAT™ is a compact, high-pressure bioreactor system that replaces the earlier VSB systems. Vigorous mixing and high oxygen transfer are critical design limitations in many conventional plants. The VERTREAT™ system has both intensive mixing and high oxygen transfers in a compact subsurface reactor, making it an efficient high-rate biological system (11–13).

3.1. Key Process Features and Advantages

The VERTREAT™ treatment system is a state-of-the-art high-rate aerobic activated sludge process. It uses an in-ground hyperbaric aeration reactor, a device that has been proven effective through more than 30 years of commercial operation. According to the manufacturer, the VERTREAT™ reactor's patented design results in a smaller reactor volume and reduced energy consumption, giving it significant capital and operating cost advantages over conventional hyperbaric aeration systems.

The following features give VERTREAT™ a strong advantage over other competing biological treatment processes (11–13):

(a) It is a proven process with similar systems already operating successfully in numerous municipal and industrial applications worldwide
(b) Operating costs are substantially lower, usually less than half that of conventional aeration processes
(c) volatile organic compounds (VOCs) emissions are minimal compared with conventional aeration processes, which can discharge up to 60% of the VOCs contained in the influent stream to atmosphere.
(d) The system is very compact and has a low space requirement, usually <20% of the space used by conventional processes
(e) There are no open aeration basins; therefore, visual impact and odor emissions are minimal
(f) The system can be economically enclosed in a building in locations where climatic conditions are unfavorable or if it is desirable for the plant to blend in architecturally with the surrounding environment
(g) The system is uncomplicated, easy to operate and maintain, and well suited to fully automated, unattended operation
(h) Concentrated waste streams with fluctuating flow rates and strengths can be treated to a high effluent quality
(i) The in-ground aeration reactor is much less likely to sustain damage in an earthquake than above-ground aeration ponds or reactors.

3.2. Process Applications

The VERTREAT™ process is ideal for treating biodegradable industrial wastewater streams and municipal wastewater. It has particular advantages in applications with the following conditions that can make conventional processes unsuitable (11–13):

(a) Sites with space constraints
(b) Wastewater streams with high VOC content
(c) Retrofits and plant expansions
(d) Applications with very concentrated wastewater streams
(e) Sites with high precipitation or extreme temperatures
(f) Sites close to residential areas
(g) Applications with fluctuating loads
(h) Locations where large unsightly plants are undesirable
(i) Sites in areas with high seismic activity
(j) Wastewater streams prone to foaming

3.3. Reactor Features

The principal difference between VERTREAT™ and other technologies employing in-ground hyperbaric aeration reactors is that the VERTREAT™ reactor has been reconfigured to incorporate three separate treatment zones (Fig. 2.9), giving it, according to the manufacturer, a significant capital and operating cost advantage over similar old-generation VSBs (11–13).

(a) *The oxidation zone* is the upper portion of the reactor and includes a central concentric draft tube for mixed liquor circulation.
(b) *The mixing zone* is immediately below the oxidation zone. Air, as required for high-rate biooxidation within the upper zone of the reactor, is injected into the mixing zone. The injected air also provides the drive mechanism for airlift circulation.
(c) *The polishing zone*, or oxygen soak zone, occupies the bottom of the reactor.

Installed by conventional drilling or excavation techniques, the VERTREAT™ reactor is typically 75 to 110 m (250 to 350 ft) deep, occupying only a fraction of the area used by conventional surface basins and using only about 10% of the air consumed by conventional aeration systems. The diameter of the reactor, nominally 0.75 to 6 m (2.5 to 20 ft), is determined by the quantity and strength of the material to be treated. A schematic of the process and

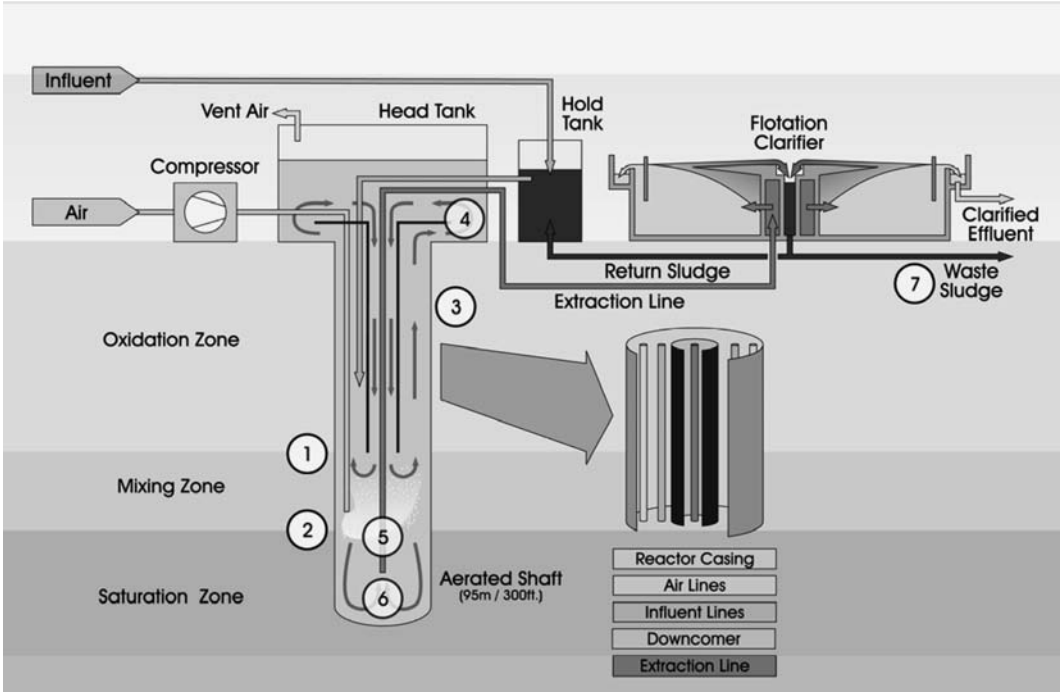

Fig. 2.9. The VERTREAT™ process (Courtesy NORAM Engineering and Construction Ltd.)

its treatment stages are included in Fig. 2.9. The seven stages in the VERTREAT™ Process are (11–13):

1st Stage—*Aeration*. Rising bubbles travel up the annulus creating a density gradient that results in airlift circulation within the oxidation zone.
2nd Stage—*Influent Injection*. Untreated influent is introduced to the recirculating liquor through the influent pipe at a level above the air injection point in the mixing zone.
3rd Stage—*Biodegradation*. High oxygen transfer rates due to the pressure and depth of injection insure high dissolved oxygen content and reaction rates within the oxidation zone.
4th Stage—*De-gassing*. Entrained spent off-gas bubbles are released to the atmosphere.
5th Stage—*Polishing*. High dissolved oxygen concentrations and residence times result in a high degree of residual BOD oxidation in this zone. Dissolved gas saturation is also utilized to drive solids separation by flotation in the clarification step that follows.
6th Stage—*Withdrawal*. Polished mixed liquor is forced from the shaft to the flotation clarifier by hydrostatic pressure.
7th Stage—*Flotation*. Rapid depressurization of the mixed liquor as it travels to the surface results in a well-aerated, low-density floc. The flotation clarifier produces a highly concentrated biomass and a high quality liquid effluent.

While space saving of the vertical bioreactor is readily apparent, the energy saving aspects may not be. The vertical reactor receives air at pressure, and requires four times the compression energy of a conventional aeration system per pound of air fed to the system (Note: at 100 psi a compressor delivers 4.7 cfm/hp, and at 7 psi a blower delivers 13 to 20 cfm/hp). However, this increased compression requirement is more than offset by the increase in oxygen transfer efficiency (OTE) in the system—approximately 70% compared to 8% OTE in conventional processes ($70/8 = 8.75{:}1$). The net result is 11% of the aeration requirement and a power consumption that is 2.2 to 2.4 times lower than that in a conventional system.

This air that is economically and efficiently introduced to the bioreactor aids in several other process functions at no incremental cost. Not only does the air satisfy the primary requirement of providing the microbes with dissolved oxygen, it serves as an air lift pump—eliminating the need for mixers in the bioreactor. Activated sludge withdrawn from the reactor is saturated with air at pressure and separates spontaneously from the mixed liquor by flotation; this flotation separation occupies a smaller area than a traditional sedimentation clarifier. In addition, the greatly reduced off-gas flow is economically treated in a small off-gas bio-filter—an achievement that would be very costly in a conventional activated sludge system due to the increased aeration requirements. Finally, if nitrification is required, the dissolved oxygen in the saturated effluent from the clarifiers carries over to the polishing nitrification biofilters, reducing the oxygen supply requirements for the blowers.

4. PROCESS THEORY AND DESIGN BASIS

4.1. Process Fundamentals

VSB treatment is a high-rate activated sludge process in which a very high mixed liquor microbial population can be maintained to achieve proportionally increased organic removal

rates. It is well known that biochemical oxidation of organic compounds is basically controlled by the following process parameters (14, 15):

(a) Concentration of organics (BOD)
(b) Concentration of active biological solids (MLVSS)
(c) Relative biodegradability of the organic mixture

Biological oxidation results in the generation of excess sludge and carbon dioxide as the primary end-products. In aerobic systems, such as those employed in the conventional activated sludge process, the respiratory or oxidative reactions provide the energy required for both synthesis and growth of biological population. Also, dissolved oxygen serves as the terminal electron acceptor, and, therefore, is essential for producing the desired end-products. In summary, the biological reactions are controlled by two basic transport mechanisms, as follows:

(a) Transport of organics (BOD)
(b) Transport of oxygen

The oxygen transport mechanism is controlled by the transfer rate of oxygen from the gas to the liquid phase, and from the liquid phase to the biological solids. When unlimited oxygen supply is available in the liquid phase, the efficiency of the biological process becomes primarily a function of the capacity of the microorganisms to assimilate organic molecules. The rate of assimilation or the rate of organics removal can be increased by increasing the MLVSS concentration and by intense mixing. Effective mixing of biological solids and organic substrate is accomplished by maintaining high liquid circulation velocities within the Vertical Shaft Bioreactor. The liquid flow velocity inside the shaft has been estimated on the order of 1 m/s (3 ft/s) with a Reynolds number greater than 100,000 (16, 17). As a result, high turbulence and intense mixing is achieved within the shaft. The driving force for liquid circulation and mixing is provided by a compressor that serves the dual function of supplying air for both liquid circulation and biological oxidation. The air supply requirements and air injection depth are generally a function of the following:

(a) Average and maximum design flow rate
(b) Strength of wastewater undergoing treatment
(c) Shaft diameter and associated friction losses due to fluid flow

The driving force for hydraulic circulation is best understood by considering the VSB start-up sequence (18). Air is injected into the riser side of the shaft. The rising bubbles, or voidage, create a density difference between the riser and downcomer causing the contents to circulate in the manner of a conventional air-lift system. The velocity of circulation increases until balanced by friction. When an equilibrium velocity is achieved (normally about 2 to 4 ft/s and certainly faster than the free-state rise rate of small air bubbles) a small quantity of air is added into the downcomer. The downward flow of the liquor drags this air to the bottom of the shaft and into the riser. The imbalance is maintained such that the overall density of liquor in the riser is always less than that in the downcomer. As equilibrium is once again established, more air is added to the downcomer injection point. This stepwise process proceeds until

eventually a dynamic equilibrium is achieved with a ratio of riser to downcomer air providing the necessary stability and oxygen demand requirements.

Because the VSB treatment process utilizes the same process concepts as the activated sludge process, the classical relationships between such process design parameters as the BOD_5 loading ratio (F/M), oxygen requirements per kg BOD_5 removed, waste sludge production (kg TSS per kg BOD_5 removed) are also applicable for the process design of the VSB.

The VSB process differs from conventional activated sludge systems in terms of equipment design and operating features. These features include the high mixed liquor suspended solids, mode, and efficiency of oxygen transfer, flow regime, and type of solids separation process. These design and operating features are summarized in Table 2.1.

4.2. Biological Properties

Principally, the VSB treatment process involves the use of aerobic metabolic capabilities for converting dissolved organics into gaseous (CO_2) and solid (waste sludge) end products. The VSB process differs from the conventional activated sludge process with respect to its flow regime, operating pressure, and oxygen tension inside the reactor. A study initiated to compare the effects of these process features on the biological properties of the sludge revealed that the waste sludge from the VSB process does not differ significantly from those experienced in conventional activated sludge systems (15). The results of this study are summarized in Table 2.2.

4.3. Oxygen Transfer

Proper design of an oxygen transfer system is essential to maintain desired minimum dissolved oxygen concentration values under both average and peak loading conditions. On a conventional activated sludge system using diffused air or mechanical surface aeration, the oxygen transfer rate is limited by the driving force (concentration differential) across the air/water interface to approximately $0.2\,kg/m^3/h$ (200 mg/L/h). As a result, the operating parameters (mixed liquor volatile suspended solids, organic loading ratio, etc.) should be carefully selected such that the oxygen demand values will not exceed the oxygen transfer capabilities.

The basic expression involved in estimating the oxygen transfer rate is (4):

$$dc/dt = K_{LA}(C_{SW} - C) \tag{2}$$

where:

dc/dt = the rate of change in dissolved oxygen concentration, $kg/m^3/h$
K_{LA} = the oxygen transfer rate coefficient, h^{-1}
C_{SW} = the oxygen saturation concentration in wastewater, mg/L
C = the minimum dissolved oxygen concentration, mg/L

According to this expression, the oxygen transfer rate in a specific waste stream or mixed liquor can be increased only by increasing the attainable saturation value (C_{SW}).

Table 2.1
Comparative design and operating criteria of VSB and activated sludge processes (4, 8)

Parameter		Air activated sludge		Oxygen activated sludge	
	Deep shaft flotation mode without primary clarification	Without primary clarification	With primary clarification	Without primary clarification	With primary clarification
Bioreactors					
Nominal detention time (h)	0.5–0.75	6–8	5–7	1.5–2.5	1.25–1.75
MLSS (mg/L)	7000–12,000	2000–3000	1500–2500	1000–6000	3500–5000
% Volatile	0.6–0.7	0.65–0.75	0.7–0.8	0.65–0.75	0.7–0.8
F/M loading (kg BOD_5/kg MLVSS · day)	0.75–1.25	0.3–0.5	0.25–0.45	0.55–0.8	0.5–0.75
Volumetric organic loading:					
(kg BOD_5/m^3day)	5.6–8.0	0.5–0.8	0.4–0.65	2.2–3.2	2.0–2.8
(lb BOD_5/day/1000 ft^3)	350–500	30–50	25–40	135–200	125–175
Sludge retention time (days)[a]	2–4	3–6	4–8	1–2	2–3
Solids Separation Unit					
Surface overflow rate:					
Average (m^3/m^2/day)	20–29	20–29	20–29	18–26	20–29
(gpd/ft^2)	500–700	500–700	500–700	450–650	500–700
Peak (m^3/m^2/day)	41–49	41–49	41–49	37–45	41–49
(gpd/ft^2)	1000–1200	1000–1200	1000–1200	900–1100	1000–1200
Mass loading (kg TSS/m^2/day)	293–439	73–12[b]	49–98	146–195	122–171
(lb TSS/day/ft^2)	60–90	15–25	10–20	30–40	25–35
Return sludge flow (% of Q)	15–25, float 30–50, bottom	25–45	25–50	30–60	30–70
Return sludge concentration (% TSS)	7–10, float 3–4, bottom	0.8–1.2	0.6–1.0	1.2–2.0	1.0–1.5
Air, Oxygen, and Power Requirements					

Table 2.1 Continued

Parameter	Deep shaft flotation mode without primary clarification	Air activated sludge		Oxygen activated sludge	
		Without primary clarification	With primary clarification	Without primary clarification	With primary clarification
Air supply rate (m^3/kg BOD_5 removed)	6–25[c]	50–9[d]			
(ft^3/lb BOD_5 removed)	100–400[c]	800–1500[d]			
Oxygen utilized (kg/kg BOD_5 removed)	2.0–2.4	0.9–1.3		1.0–1.4	
Oxygen transfer efficiency in wastewater, % (O_2 utilized/O_2 supplied)	40–90[c]	8–15[d]		90–95	
Oxygen transfer rate in wastewater:					
(kg O_2/wire kWh)	0.9–2.7[c]	0.9–1.5[d]		1.2–1.5[e]	
(lb O_2/wire hp-h)	1.5–4.5[c]	1.5–2.5[d]		2.0–2.5[e]	
Aeration system power requirement:					
Wire (kWh/1000 m^3)	80–280	112–177	64–112	121–145	72–88
Wire (hp-h/MG)	500–1800	700–1200	400–700	750–900	450–550
Sludge Production					
Primary sludge TSS[f] (g/m^3)	–	–	132	–	132
(lb/MG)	–	–	1100	–	1100
Waste activated sludge:					
VSS (g/m^3)	72–90	96–132	54–78	96–120	54–66
(lb/MG)	600–750	800–1100	450–650	800–1000	450–550
(kg/kg BOD_5 removed)	0.4–0.5	0.55–0.75	0.5–0.7	0.55–0.65	0.5–0.6
TSS (g/m^3)	108–138	138–186	72–102	138–168	72–90
(lb/MG)	900–1150	1150–1550	600–850	1150–1400	600–750
(kg/kg BOD_5 removed)	0.6–0.75	0.75–1.05	0.65–0.95	0.75–0.95	0.65–0.8

Table 2.1 Continued

Parameter	Deep shaft flotation mode without primary clarification	Air activated sludge		Oxygen activated sludge	
		Without primary clarification	With primary clarification	Without primary clarification	With primary clarification
Total plant raw and waste sludge TSS:					
(g/m^3)	108–138	138–186	204–234	138–168	204–222
(lb/MG)	900–1150	1150–1550	1700–1950	1150–1400	1700–1850
Final effluent solids:[g]					
(g/m^3)	13	14	16	14	16
(lb/MG)	110	120	130	120	130
TSS (g/m^3)	20	20	20	20	20
(lb/MG)	170	170	170	170	170
Total sludge production (waste and effluent) in secondary system:					
VSS (g/m^3)	85–103	110–146	70–94	110–134	70–82
(lb/MG)	710–860	920–1220	580–700	920–1120	560–680
(kg/kg BOD_5 removed)	0.45–0.55	0.6–0.8	0.65–0.85	0.6–0.75	0.65–0.75
TSS (g/m^3)	128–158	158–206	92–122	158–188	92–110
(lb/MG)	1070–1320	1320–1720	770–1020	1320–1570	770–920
(kg/kg BOD_5 removed)	0.7–0.9	0.9–1.15	0.85–1.1	0.9–1.05	0.85–1.0

[a] Defined as kg MLSS in bioreactor/(kg TSS lost in waste activated sludge and final effluent/day).
[b] Dry weather peak.
[c] Depends on shaft diameter and degree of air tuning in shaft (Eco-II vs. Eco-III).
[d] Lower values representative of coarse bubble diffusers: higher values representative of fine bubble diffusers.
[e] Lower values apply to systems employing pressure swing adsorption (PSA) oxygen generation: higher values apply to systems employing cryogenic oxygen generation.
[f] Calculated on the basis of 65% removal of an assumed raw wastewater concentration of 200 mg/L.
[g] Assumes a final effluent TSS of 20 mg/L and final effluent solids volatility in accordance with assumed mixed liquor volatility of respective processes.

Table 2.2
Comparison between VSB and conventional activated sludge (4, 5)

Sludge properties[a] (or) components	Deep shaft reactor	Conventional activated sludge
ATP content (mg/L)	0.806	0.537–0.991
Specific oxygen uptake rate (g/kg.h)	40	14.5–57.7
Michaelis-Menton[b] growth constant—K_s (mg/L)	50	20–50
Physical Characteristics[c]		
Specific resistance (m/kg $\times 10^{14}$)	1.29	8.54[c]
Compressibility index	0.85	0.78
Waste sludge concentration (%)	2.1	0.94

[a] Average values for sludge properties are reported; comparison was made of sludges produced from the treatment of primarily domestic wastewaters.

[b] Refers to the concentration of BOD_5 (raw wastewater) at which the specific oxygen uptake rate is one-half the maximum value. The term "Michaelis-Menton Growth Constant" is used for comparison of specific oxygen uptake rate values because of the belief that the theory of enzyme reaction kinetics is directly applicable in describing the growth or BOD_5 removal kinetics in the activated sludge process.

[c] Physical characteristics for waste activated sludge from conventional air activated sludge was determined utilizing aerobically digested sludge samples.

According to Henry's law, the saturation value can be increased by raising the partial pressure of the gas requiring dissolution. This can be accomplished by either of the following methods:

(a) Increasing the mole concentration of oxygen in the source (enriched oxygen systems) such as those used in the pure oxygen activated sludge process
(b) Increasing the system operating pressure as in the case of the VSB

In a pure oxygen activated sludge system, the oxygen transfer rates are approximately five times greater than in systems using air diffusion or mechanical surface aeration. In the case of a VSB, the operating pressures are increased to 1,520 kPa (15 atm) and, therefore, the oxygen transfer rate is similarly increased to 2,000 to 3,000 mg/L/h (150 to 200 lb/1,000 ft^3/h). The aeration bubble contact time in the vertical bio-reactor is in the order of 3 to 4 min rather than 10 to 15 s in a shallow surface basin (19). The increased oxygenation capacity allows the system to operate with higher mixed liquor suspended solids concentrations and, therefore, with lower aeration periods than in the conventional activated sludge process.

4.4. Organic Loading

A relationship was developed between organic loading ratio (F/M) and the oxygen transfer requirement for various MLVSSs. This relationship is illustrated graphically in Fig. 2.10, which indicates that there is a limiting loading ratio (F/M) for each mixed liquor suspended solids concentration, above which the oxygen demand requirements cannot be satisfied by conventional methods. For illustrative purposes, the upper limit for oxygenation capacity has been assumed at 0.08 kg/m^3/h (5 lb/1000 ft^3/h) of aeration volume for conventional air

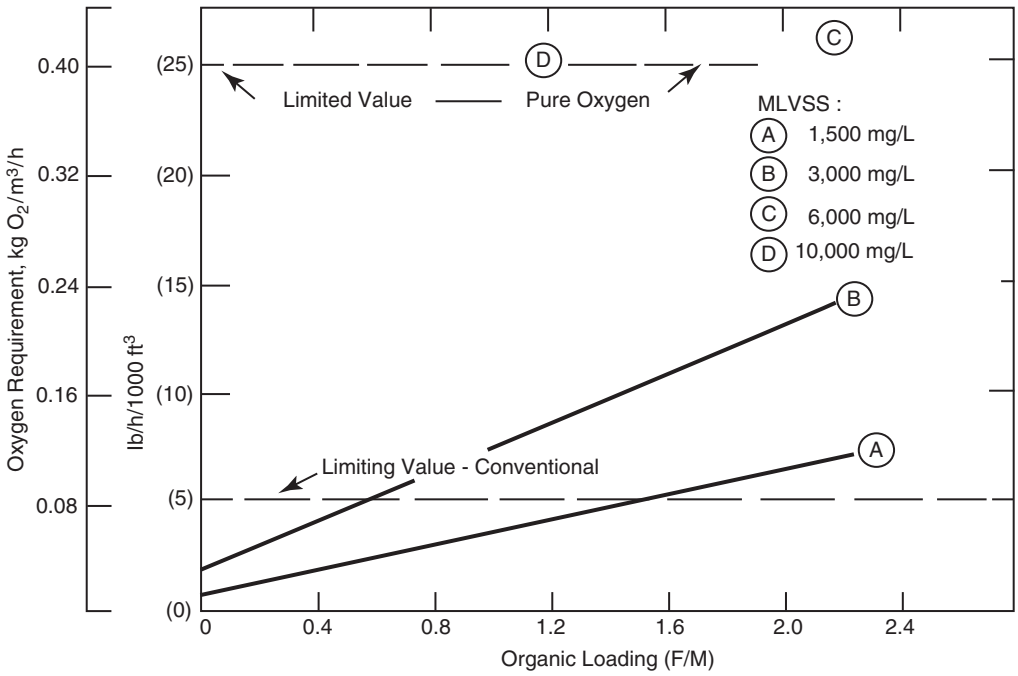

Fig. 2.10. Relationship between organic loading (F/M) and oxygen requirement (4)

systems. For example, the organic loading ratio (F/M) must be maintained below 0.55 when the operating MLVSS is 3000 mg/L in order that the aeration capabilities of the conventional equipment will not be exceeded. By reiteration of this technique, a limiting envelope was developed that relates the organic loading ratio (FM), MLVSS, and oxygenation capacity (4).

Similarly, another limiting envelope was developed for pure oxygen systems with a maximum oxygenation capacity assumed at 0.40 $kg/m^3/h$ (25 $lb/1{,}000\ ft^3/h$). Figure 2.11 shows these limiting envelopes for the conventional and enriched oxygen systems as well as the operating zones for conventional, enriched oxygen and VSB systems. It is to be recognized that these limiting curves are developed with assumed or preselected values for oxygenation capacities. Actual limiting values may differ depending on the aeration device selected for a particular application (e.g., fine bubble, coarse bubble, aeration basin depth, mechanical surface aeration). In addition, the limiting envelopes for the different technologies may overlap.

It is evident from this process evaluation that one of the major constraints imposed on the design of an aerobic biological treatment process is the capability of the aeration equipment to maintain an aerobic environment. The VSB process is capable of exceeding these limits as the system can achieve up to 75% or even 90% oxygen transfer efficiency. As a result, organic loading ratios (F/M) as high as 2.0 can be used with mixed liquor volatile suspended solids concentration values of up to 10,000 mg/L, thereby reducing the aeration periods to 30 min or less (4).

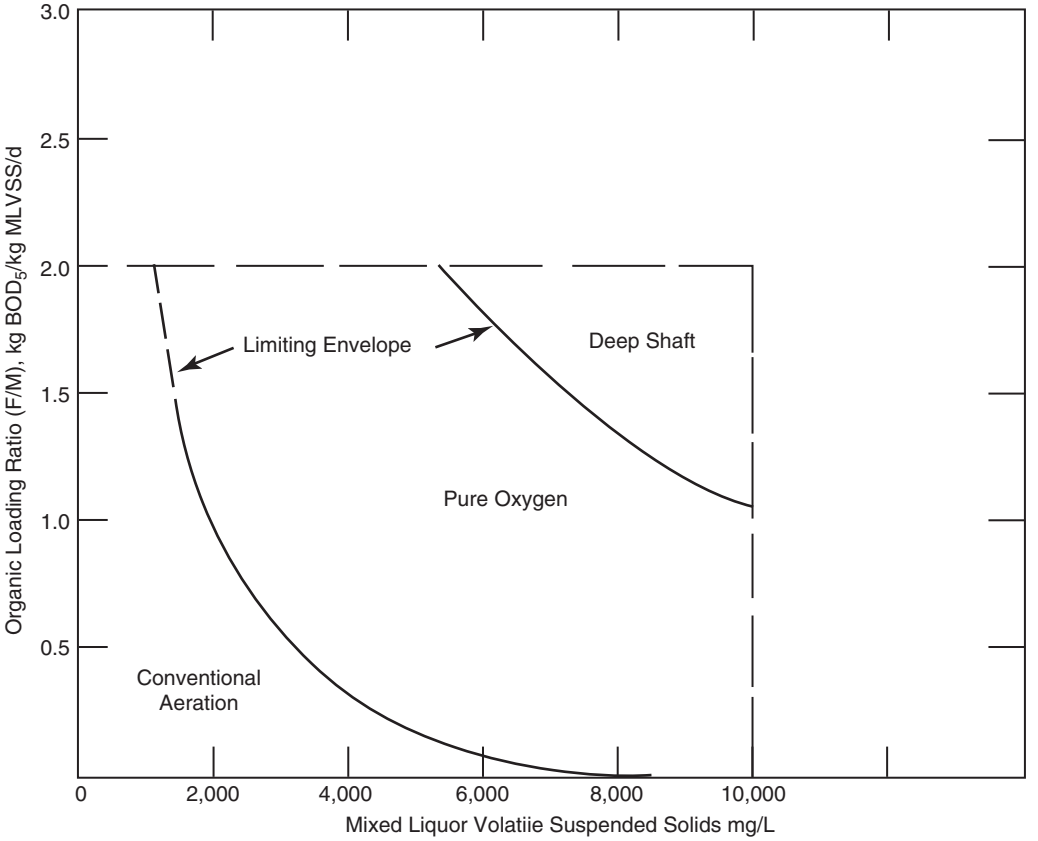

Fig. 2.11. Comparison of oxygen transfer limiting envelops for biological treatment processes (4)

4.5. Solids Separation

One of the major considerations in the design of an aerobic biological wastewater treatment system involves the incorporation of an effective solids separation process unit. Gravity sedimentation units have served this purpose reasonably well within the operating range for conventional systems (MLSS between 2,000 and 3,000 mg/L) and oxygen-enriched systems (MLSS between 4,000 and 6,000 mg/L). These units serve the dual purpose of producing a clarified effluent and a source of sludge for recirculation. This latter function is critical in maintaining the biological integrity of the aeration basin to produce a flocculant biomass that can readily settle. Extensive studies on the gravity settling and thickening characteristics of activated sludge have indicated that the process is effective when the suspended solids concentration values are maintained below 6,000 mg/L (9, 10). This will permit operating the gravity sedimentation units with a reasonable sludge blanket depth (0.25 to 1 m) and within a recommended solids flux of 29 to 120 $kg/m^2/d$ (6 to 25 $lb/ft^2/d$).

Table 2.3
Design and operating parameters for VSB solids separation processes (4, 20)

Parameter	Flotation mode	Sedimentation mode
Hydraulic overflow rate ($m^3/m^2/d$ (gpd/ft^2))	20 (500)	10 (250)
Mass loading ($kg/m^2/d$ (lb/d/ft^2))	320 (66)	103 (21)
Float solids concentration (%)	7–10	ND
Sink solids concentration (%)	3–4	1–2

ND: No Data.

Earlier versions of the VSB process recognized these limitations and the process was designed to operate with suspended solids concentration values between 5000 and 6000 mg/L. However, the North American versions of the VSB process have adopted a dissolved air flotation process as the terminal unit operation, and utilize the available dissolved gases. The dissolved gases present in the VSB simulate the pressure vessel in the dissolved air flotation process, and provide the driving force during solids separation. The process possesses additional advantages in producing a significantly higher float solids concentration (4% to 7%) than the underflow solids concentration from a typical gravity sedimentation unit (1% to 3%). Table 2.3 summarizes the design and operating features of the two concepts (20).

5. VARIATIONS OF THE BASIC VSB

The vertical shaft bio-reactor can be designed to create the environment that is optimal for the desired biological process. For instance, in order to maximize cell production, a multiple port feed injection is required. This minimizes the period of time for cell digestion of organic carbon starvation. Additionally, research indicates that substrate to cell conversion is reduced by cycling DO concentrations. This is due to a disruption of the energy balance in the cell, which interrupts the cell synthesis step in favor of converting more of the carbon to CO_2. Conversely, to minimize biomass production, such as in waste water treatment, a longer period of carbon limitation is required. In other applications of wastewater treatment, the reactor can be designed for nitrogen and phosphorus removal by creating the appropriate aerobic, anoxic, and anaerobic zones. In the biological nutrient removal (BNR) application, the bio-reactor is configured to provide successive zones that favor the preferred type of biological activity similar to the function of an oxidation ditch but oriented vertically instead of horizontally. The following subsections give is a brief description of the various variations (19).

5.1. Single Zone Vertical Shaft Bioreactors

The basic vertical shaft bio-reactor comprises an outer casing and an inner tube extending to within 2 to 3 m (6 to 10 ft) of the bottom of the reactor. The inner tube is a down-flow conduit while the annular space between the downflow tube and the casing is the upflow passageway. Air can be injected in both the downflow stream and the upflow stream or only in the upflow stream. The basic design vertical bioreactor provides a rapid removal of organic contaminants (BOD), with low energy cost and very small foot-print. This design is suitable for treatment

of wastewater where the effluent criteria requires <90% to 95% BOD_5 removal and total suspended solids are in the order of 25 to 75 mg/L. A typical application would be a roughing plant discharging to a sewer collection system or to an on site polishing plant. In this design the organic loading can be in the order of 8 kg BOD_5/m^3 (0.5 lb BOD_5/ft^3). Sludge production can be expected to be in the order of 0.5 to 0.6 kg solids/kg BOD_5 removed because at these loadings, there is virtually no zone of carbon starvation in the reactor. However, the effect high energy mixing continue to function in the vertical bioreactor thus producing less sludge than a conventional high rate system operating under similar conditions. Solids separation and dewatering are expected to be rather difficult.

5.2. Multi-Zone Vertical Shaft Bioreactors

When higher quality effluents are required, the organic loading on the basic design bioreactor must be lowered or; the bioreactor can be reconfigured to incorporate an internal polishing zone. In the basic reactor, the entire re-circulating flow is treated down to effluent endogenous respiration rate (about 4 to 6 mg O_2/g TSS/h) that is equivalent to 2 to 3 mg/L of filtered effluent BOD_5). In the multi zone reactor, a separate polishing zone is provided either above or below the circulating zone. The circulating zone in the reactor comprises the head tank, the down flow tube and the annular space between the down flow tube and the casing. The polishing zone receives only the effluent portion of the circulating flow and treats only that portion down to endogenous respiration rates. Typically the effluent flow is 2% to 20% of the circulating flow, depending on the waste strength, and therefore a relatively small volume of the reactor is required for polishing treatment. The multi zone reactor is about twice as efficient as the basic reactor design.

The polishing zone is positioned at the bottom of the reactor when solids separation is by flotation clarification and at the top of the reactor when solids separation is by sedimentation clarification. The basic design and the multi-zone reactor design take advantage of the generally accepted notion of 'first order' rate of substrate oxidation (actually, in a vertical bio-reactor the kinetic rate of reaction is closer to 0.6-order rate). That is to say, when essentially complete treatment of the readily biodegradable substrate is required, about 80% of the BOD_5 is removed in 20% to 30% of the calculated retention time while substantially all of the remaining 20% of the BOD_5 is removed in the remaining 70% to 80% of the time. In the basic reactor design case, these long residence times lead to low organic loading rates in the entire reactor. This detracts from the high oxygen transfer capability of the reactor and reduces the overall efficiency. However in the multi zone reactor design, by routing only the effluent flow through the polishing zone, and not the entire circulating flow, the required residence time in the polishing zone can be achieved in 2% to 20% of the reactor volume.

5.3. Multi-channel Vertical Shaft Bioreactors

When refractory compounds are known to exist in the waste water the VSB system can be designed with two reactors, a smaller reactor inside a larger reactor. The larger reactor treats substantially all the readily biodegradable BOD. A portion of the concentrated return activated sludge (RAS) from the clarifier is directed to the smaller reactor. The dissolved air flotation separation process results in a high percentage of refractory and surface active compounds

being adsorbed onto or trapped in the return sludge. The concentrated RAS stream, or a portion thereof, is directed to the smaller shaft for stabilization. Typically, a residence time of 4 to 6 hours for brewery wastewater and 24 to 48 hours for refinery wastewater is required to biodegrade the adsorbed refractory compounds. After partial digestion in the small reactor, the RAS is recycled to the big reactor or sent to wasted sludge tank. The operating principle of this smaller reactor is similar to that of a contact stabilization process. This type of multi-channel reactor produces about 20% to 40% less sludge than the basic reactor configuration.

5.4. Multi-Stage Vertical Shaft Bioreactors

When biological nutrient removal (BNR) is required, the reactor is reconfigured to provide a nitrification step/polishing zone, a denitrification step/feed tank arrangement, and an anaerobic step (either internal or external to the reactor) for the production of volatile fatty acid (VFA). The nitrification step consumes about 30% of the oxygen required for BOD removal and requires a source of inorganic carbon (alkalinity) preferably in the form of carbonic acid or carbonate. A convenient method of satisfying these conditions in a vertical bioreactor is to aerate the nitrification zone with the reactor off gas that contains typically about 45% oxygen and 16% to 17% CO_2 for a domestic wastewater application. Since the bioreactor off-gas is under pressure, no air blowers are required. Surplus sludge from the nitrification zone is returned, along with a small denitrified recycle stream, to the bio-oxidation step.

5.5. Thermophilic Vertical Shaft Bioreactors

For some very high strength industrial wastes the vertical shaft bioreactor can be operated in the thermophilic range of 50°C to 70°C. Typically the change in temperature across the reactor is about 3°C/1,000 mg/L of BOD_5. The advantages of thermophilic operation are:

(a) Low sludge production
(b) Low nutrient consumption
(c) Higher rates of bio-oxidation

Typically, industrial waste water is alkaline (e.g., brewery, refinery, dairy) and the CO_2 concentration from the bio-oxidation process in the bioreactor is sufficient to neutralize influent pH levels of 10 and higher. The disadvantages of thermophilic operation are:

(a) Poorer oxygen transfer due to the high temperature
(b) Thermophilic bacteria are less robust than the mesophilic bacteria
(c) Bio-diversity is less and because dispersed thermophilic bacteria dominate the biomass, the solids separation is more difficult

6. PROCESS DESIGN CONSIDERATIONS

The activated sludge process requires relatively large amounts of energy to transfer adequate amounts of oxygen for carrying out the biological reactions. When these demands exceed the oxygen transfer capabilities of conventional equipment (diffused air or mechanical surface aeration), the aeration basin volume is generally increased to balance the oxygen demand–supply characteristics. As a result, the design and operating characteristics of conventional systems are often dictated by the limitations imposed by oxygen transfer equipment.

Vertical Shaft bioreactors are designed to operate with 90 to 150 m (300 to 500 ft) of hydrostatic pressure with oxygenation capacities between 2,000 and 3,000 mg/L/h. As a result, the design of Vertical Shaft bioreactors is basically dependent on the organic removal rate and the availability of a consistent source of recycle biomass. In general, the design of a biological reactor involves consideration of the following (4):

(a) Providing adequate mixing to maintain mixed liquor solids in suspension, and to improve the opportunity for contact between biological solids and organics
(b) Providing adequate residence time in the reactor for achieving the desired removal efficiency
(c) Providing adequate facilities for recycling sludge and for maintaining the desired mixed liquor volatile suspended solids concentration

Mixing in VSB is accomplished by maintaining sufficient velocities and turbulence through the shaft (1 to 2 m/s). During startup, the flow inside the VSB is initiated by injecting air into the riser section. The differential hydrostatic head, developed due to the voidage difference between the downcomer and the riser sections, is adequate to initiate and maintain flow through the shaft. The driving force (F) required to maintain flow through the reactor is estimated from the voidage head difference and the friction loss, as follows:

$$F = \text{voidage head} - \text{friction loss} \tag{3}$$

In general, the voidage head difference is adjusted by controlling the air injection depth to the downcomer. The air requirements and the air injection depth are usually selected to maintain forward flow under all conditions (average and peak flow conditions). For domestic wastewaters ($BOD_5 = 200$ mg/L), the air flow requirements are primarily dictated by the required driving force to maintain flow. In the case of high strength wastewaters, the air flow requirements may be dictated by the wastewater's organic strength and oxygen requirements (4).

The residence time and extremely high pressure (up to 1,520 kPa; 15 atm) available in the lower sections of the Vertical Shaft Bioreactor are sufficient to achieve nearly complete dissolution of oxygen. For design purposes, it is usually assumed that 90% of the oxygen supply goes into solution during passage through the reactor. This is equivalent to 0.25-kg oxygen for each cubic meter of air injected into the reactor. The total air requirements for biological oxidation can thus be estimated from the raw wastewater characteristics and treatment requirements.

Optimization studies conducted with air diffusion in Vertical Shaft Bioreactors indicate that, at 90% oxygen absorption efficiency, oxygen demand rates of up to 1 kg/m^3/h can be satisfied with a 135 m (450 ft) deep reactor. In general, an operating depth of between 100 and 150 m (328 to 492 ft) is usually selected for design of the VSB, taking into consideration the patent regulations on other similar processes (21). Figure 2.12 shows the dissolved oxygen and BOD profiles normally anticipated inside reactor systems. The design and operation criteria were presented in Table 2.1.

VSBs have the same process concepts and capabilities as conventional activated sludge systems. Because of the high mixed liquor volatile solids maintained in the VSB, volumetric organic removal rates are higher than in the equivalent conventional concept. As a result, the

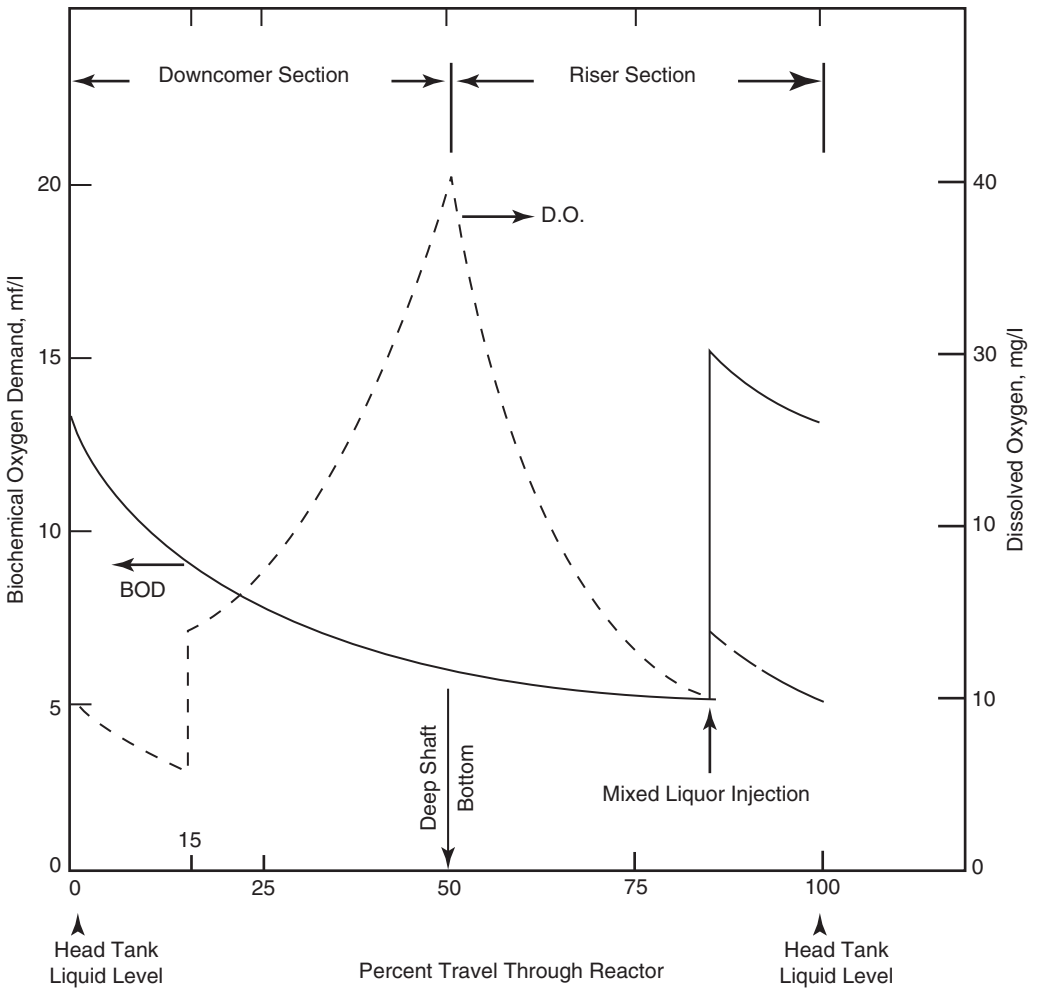

Fig. 2.12. Dissolved oxygen and BOD profiles for VSB (4)

aeration period is relatively low and is on the order of 30 to 60 min. Based on an average flow-through velocity of 1 m/s (3.05 ft/s) inside the VSB, the average turnover rate for the mixed liquor is approximately once every 5 minutes when the reactor depth is 150 m (457 ft) (22). This circulating turbulent mixed liquor serves as the dilution medium for the influent waste stream to the reactor. The dilution factor is a function of the mean residence time (t) of the influent waste stream in the reactor and the flow-through velocity inside (v). The dilution factor can be expressed as follows:

$$Q_{\mathrm{i}}/Q_{\mathrm{R}} = (H/v)/t \tag{4}$$

where:

Q_i = influent waste flow rate, m^3/h
Q_R = mixed liquor flow rate through the VSB, m^3/h
H = depth of VSB, m
v = flow-through velocity inside the reactor, m/h
t = mean residence in the reactor, h

This design feature of the VSB aids in minimizing the effects of shock loads on system performance.

Even though the flow pattern inside the reactor resembles plug flow for each passage, the mixed liquor turnover rate and the external dilution aid the system to approach complete-mix status, and therefore the system is relatively stable to variations in influent characteristics. Figure 2.13 shows the comparison in concentration profile within completely mixed, plug flow, and VSB.

Because of the ability of the VSB to achieve oxygen transfer efficiencies of up to 90%, the system is suitable for the joint treatment of high-strength industrial and municipal wastewaters. Similarly, the system is also suitable for pretreatment of industrial wastewaters (23–25).

Because of the relatively low residence time utilized in the design of the VSB, the system is susceptible to upsets due to sustained hydraulic peak flows. The reactor is usually equipped with a two-speed drive mechanism for the float skimmer for adjusting the recycle sludge flow rate. For the same reason, it is essential to adequately define the average and maximum flow conditions during the design of a VSB.

7. OPERATION AND MAINTENANCE CONSIDERATIONS

The VSB is very simple in configuration and has no moving parts inside the shaft. As a result, the requirement for maintenance of the shaft components themselves is minimal and is less than what is required for conventional activated sludge processes equipped with air diffusers. The high pressure (790 kPa; 100 psi) compressors used in the VSB process, however, will require increased maintenance as compared to the low pressure blowers (<79 kPa) or mechanical surface aerators used in conventional systems (24). Similarly, the operation of the dissolved air flotation process will require additional training and increased operator monitoring as compared to a gravity sedimentation process.

One Eco version of the VSB has eliminated most automatic instrumentation and controls, thereby making it less complicated than conventional processes. This is especially true with respect to the sludge recirculation system that is set at a constant rate during normal flow conditions.

Because the VSBs are installed subsurface, the mixed liquor inside the reactor is not subject to wide seasonal variations in temperature. Therefore, process operating parameters can be maintained at a steady rate year-round and less operator attention will be required.

A disadvantage of the VSB process, however, is the inability to visually observe mixed liquor contents so that process upsets can be detected immediately (4). In general, the VSB process is not appreciably different from conventional activated sludge systems, and it is not

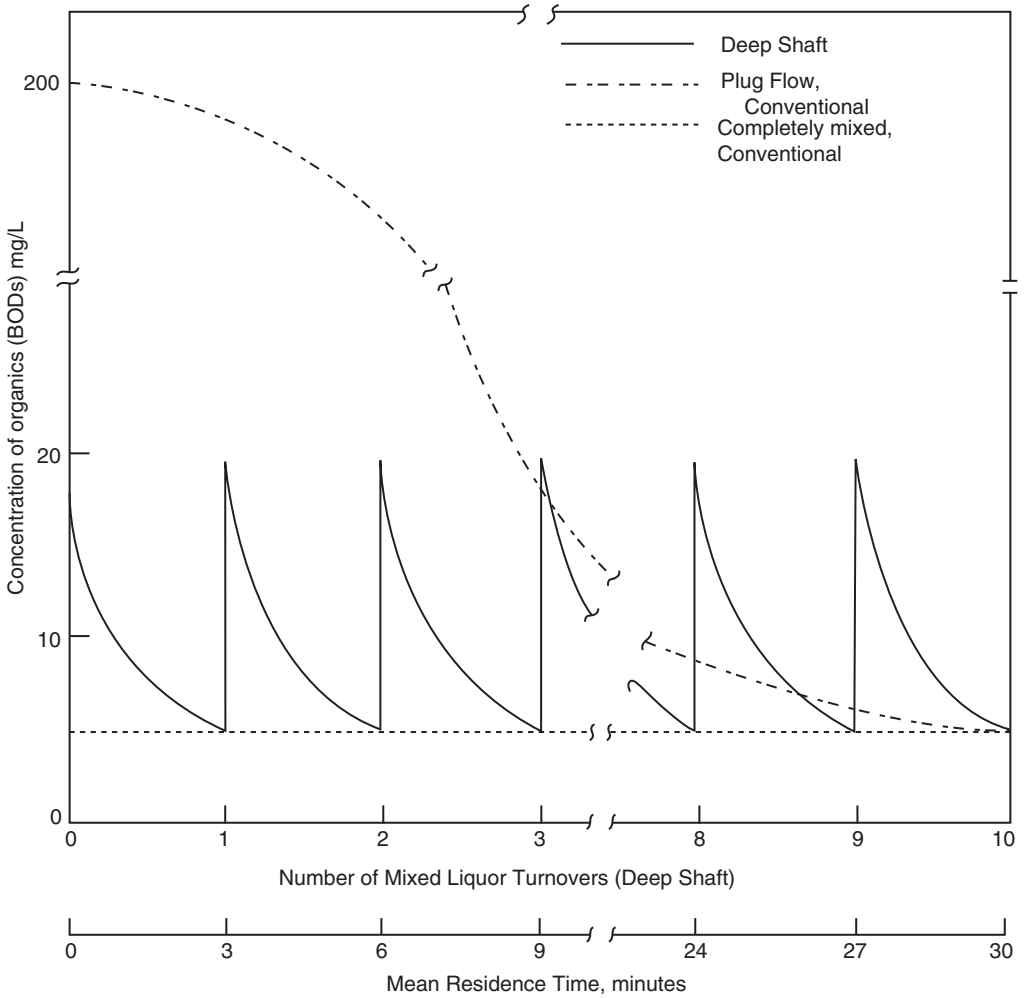

Fig. 2.13. Comparison of BOD concentration profiles for conventional and VSB systems (4)

expected to require any specialized skills. Therefore, the staffing requirements will be similar to the conventional systems of equivalent size. Because of this similarity, the VSB process may be suitable for expanding existing activated sludge plants where space restrictions prevail.

8. COMPARISON WITH EQUIVALENT TECHNOLOGY

8.1. Equivalent Conventional Concept

The VSB treatment system is a high rate activated sludge process in which the shallow aeration basins of 3 to 10 m (9 to 30 ft) are replaced with deep subsurface reactors of 90 to 250 m (270 to 760 ft). In addition, the North American version of the VSB process utilizes dissolved air flotation for final clarification of mixed liquor suspended solids. According to

an EPA report (4), the most suitable equivalent technology for comparison purposes is the enriched oxygen process (pure oxygen). Aside from other similarities, the pure oxygen system is usually designed to operate with high mixed liquor suspended solids (4,000 to 6,000 mg/L) and with a high dissolved oxygen concentration (5 to 7 mg/L). These design features allow the bioreactor to operate under high organic loadings (F/M) end with reduced aeration volume similar to those achievable in the VSB process.

Other similarities between the pure oxygen activated sludge and VSB alternative include the high oxygen tension within the bioreactor and the resultant low waste sludge generation. A comparative analysis of these design features and operating criteria were presented in Table 2.1. This comparison indicates that design criteria such as the nominal detention time, mixed liquor suspended solids (MLSS), organic loading, and sludge age are within the same range for the oxygen-activated sludge and the VSB process. In general, the comparative analysis of design and operating criteria indicates that the two processes are similar except for the oxygen utilization efficiency and the return sludge concentration values. For these reasons, the pure oxygen system is selected as the equivalent technology for comparison with the VSB process. The air-activated sludge process is included in the evaluation in order to establish a baseline technology in the comparative analysis. For the 1892 m^3/d (0.5 MGD) facility, conventional activated sludge was used as the baseline technology, whereas high-rate activated sludge was used as the baseline technology for the 16,925 m^3/d (5.0 MGD) and 37,850 m^3/d (10.0 MGD) facilities (4).

8.2. Land Area

One of the significant advantages of the VSB system is the reduced land area requirement as compared to the conventional air or pure oxygen activated sludge systems (4, 12, 13, 26). This feature makes the VSB system especially attractive for consideration in land restricted areas, and in expanding existing facilities where land availability is limited. Figure 2.14 shows the relative land area requirements for the VSB and conventional air-activated sludge systems (4). Based on the design criteria presented in Table 2.1, it is likely that the land area requirements for the pure oxygen-activated sludge system will be similar to the conventional air-activated sludge process. This is due to the fact that any space reductions realized in aeration tank sizing will be compensated by the additional area required for installing oxygen supply equipment.

8.3. Cost

Based on experience with the VSB process, the major cost element is associated with the installation of the reactor itself. The fixed cost associated with well drilling and shaft installation, including electrical, mechanical, and instrumentation devices, has been estimated to be between 30% and 50% of the total project cost (27–29) The cost of drilling is subject to variation depending on geological conditions, the availability of drilling rigs, and their demand for other more competitive purposes (e.g., oil well drilling).

For industrial wastewater treatment, the capital cost of a VSB system is lower than that in conventional plants of similar size. Decreased land requirements, considerably less surface tanks (less concrete) and fewer pumps are some of the key elements decreasing the capital cost (13).

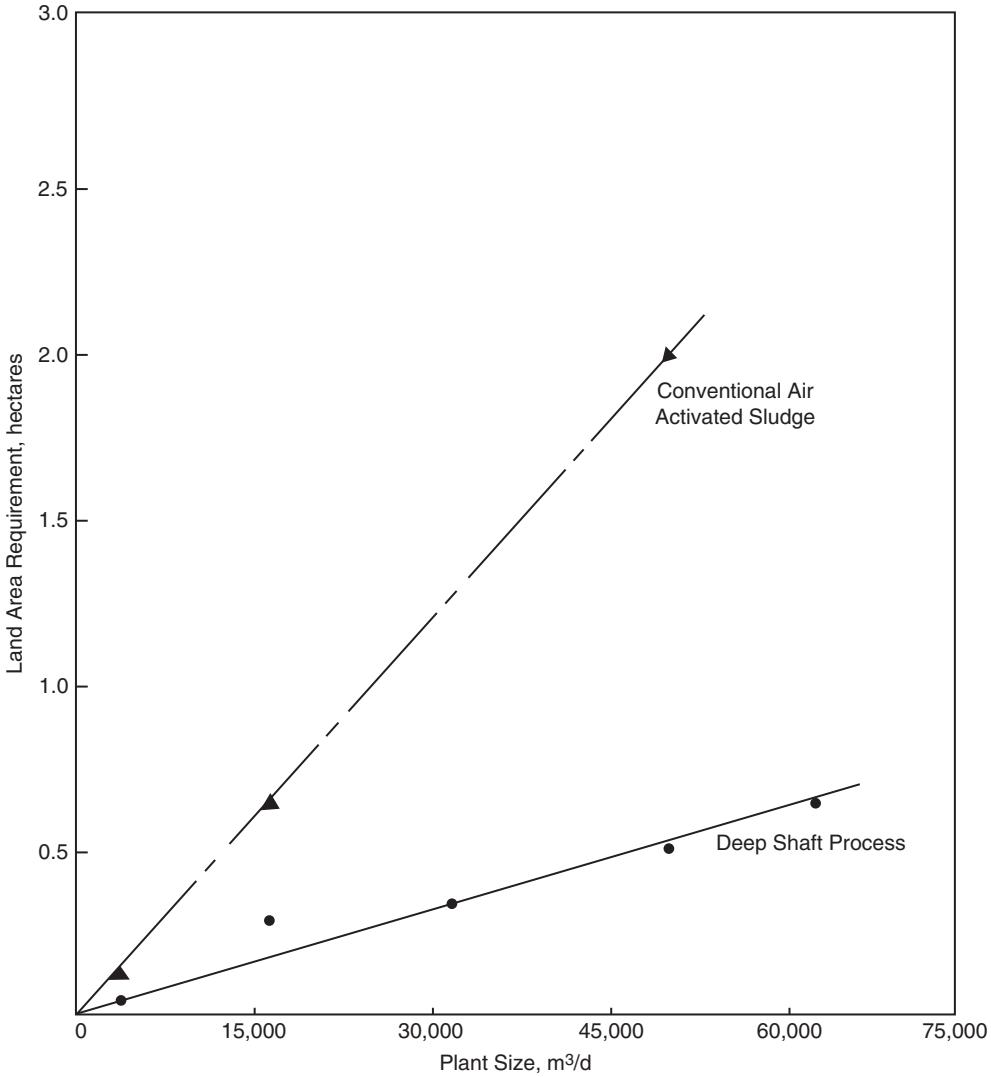

Fig. 2.14. Land area requirements for conventional activated sludge and VSB processes (4)

Several factors support the reduced capital costs and land requirements of VSB systems. These factors account for their requiring 20% of the total land required for conventional plants of equivalent capacity—reducing visual and environmental impact. Some of these factors include (13):

(a) Eighty percent of the bioreactor volume is below grade—eliminating large surface tanks.
(b) Due to the high oxygen transfer efficiency, the residence time required in the bioreactor is decreased relative to conventional technologies—making the required reactor volume smaller.

(c) The solids are easily float-thickened to 4% solids concentration. Float thickening in this manner significantly reduces the size of the clarification system and the downstream dewatering facility.

The most significant savings realized in the VSB process relate to the aeration system (13). The basis of the process is that the oxygen transfer efficiency is significantly higher than that in a conventional aerobic biological treatment system due to the pressure at the depth where air is introduced to the bioreactor. The oxygen transfer efficiency exceeds 75% vs. conventional air activated sludge facilities that can achieve only 8% to 10% oxygen transfer efficiency. Air that is economically and efficiently introduced to the bioreactor aids in several other process functions at no incremental cost. Not only does the air satisfy the primary requirement of providing the microbes with dissolved oxygen, it serves as an air lift pump—eliminating the need for mixers in the bioreactor. Air indirectly provides the dissolved gases necessary for solids flotation in the flotation clarifier that follows the bioreactor—decreasing the size of the downstream separation equipment. The highly efficient aeration system allows for low power consumption and low chemical usage. This amounts to low direct variable operating costs in the system—approximately USD 3.27/100 lb BOD_5 destroyed compared to direct variable operating costs in conventional systems that are approximately USD 8.70/100 lb BOD_5 destroyed (13).

Economic analysis of three technologies for the treatment of municipal wastewater was considered by a US, EPA report (4). The initial investment cost (capital cost), the annual operation and maintenance cost, and the present worth cost of the total treatment systems were evaluated. The cost estimates developed by the US EPA for evaluating innovative and alternative technologies were used as the primary source for estimating installed capital and annual operation and maintenance costs for the pure oxygen and conventional activated sludge processes. These cost estimates were supplemented with cost figures from the Area-wide Assessment Procedures manual to include structural and nonstructural cost components (e.g., influent pumping or lift station, and miscellaneous structures such as control and operations buildings, outfall sewer) (30).

The VSB portion of the treatment plants included the vertical shaft bioreactor(s), flotation separator units, and the control building for these components. These cost estimates were supplemented with estimates for remaining process units (e.g., sludge handling and treatment, preliminary treatment, disinfection, influent, and effluent structures) utilizing the same cost curves as the equivalent and baseline technology alternatives. All cost estimates were updated from 1980 to reflect 2008 construction costs using the Cost Index for Utilities (Appendix); all costs were multiplied by a factor of $522.16/277.60 = 1.99$ (31).

The VSB process showed between 26% and 33% savings in installed capital costs over the pure oxygen activated sludge system for the treatment of municipal flows ranges for which the comparative analysis was prepared. For the treatment of highly concentrated industrial wastewaters, the VSB process will in most cases be even much more outstanding and competitive than the conventional activated sludge processes.

8.4. Energy

The major energy requirement in biological wastewater treatment systems is the biological reactor in which the oxygen demand requirements must be supplied from external sources.

The VSB process is no exception to this requirement, since the oxygen is supplied using high pressure compressors with discharge pressures of 790 kPa (100 psi). The actual energy requirements for a vertical shaft bioreactor are governed by the following (4):

(a) Organic and hydraulic load for average and peak conditions
(b) Mixed liquor volatile suspended solids (MLVSS)
(c) Air requirements for liquid circulation
(d) Shaft diameter

In general, shafts smaller than 1 m (3 ft) in diameter may require supplemental air to maintain mixed liquor circulating velocities in treating normal strength domestic wastewater (25). When optimum organic loading conditions prevail, oxygen transfer efficiencies up to 6 kg O_2/kWh (9.8 lb O_2/hp) can be realized (6). On the other hand, small diameter shafts treating weak wastewaters can realize power economies in the range between 2 and 3 kg O_2/kWh (3.3 to 4.9 lb O_2/hp).

An approach similar to that utilized for the cost comparison was used for estimating the energy requirements for various size plants (4). It is evident that the VSB process benefits (cost and energy) can be more outstanding when the raw wastewater strength is greater than normal domestic wastewater. This is because the energy requirements for the VSB process treating domestic wastewaters are based on the requirement for maintaining liquid circulation velocities rather than on the basis of BOD_5 removal. When the raw wastewater BOD_5 concentration is high, the cost and energy savings are more likely to be in favor of the VSB (4).

9. CASE STUDIES

In this Section the successful application of the VSB process for the treatment of two types of industrial wastewater and one municipal wastewater is discussed:

(a) Dairy plant wastewater treatment
(b) Refinery wastewater treatment
(c) Municipal Wastewater Treatment

9.1. Dairy Plant Wastewater Treatment

9.1.1. Process Description

The wastewater treatment plant for the dairy effluent consists of 3.2 ML (850,000 gal) of equalization capacity, two 2.75 m (9 ft) diameter 94 m (308 ft) deep VERTREAT™ VSBs, and two rectangular flotation clarifiers (11, 13). Influent is pumped through basket strainers and coolers to the bioreactors. It is then cooled to maintain mesophilic conditions in the bioreactors (25°C to 40°C). Provision has been made for nutrient addition and pH adjustment of the influent. The dissolved oxygen concentration in the VSBs is used to control the aeration from four 450-scfm compressors. Vent air containing hydrogen sulfide is extracted from the equalization tank with a blower, and is injected in the bioreactor for bio-oxidation. Polished mixed liquor is withdrawn from the bioreactor soak zones to two flotation clarifiers where the biomass is removed from the treated effluent. A portion of the separated sludge is returned to

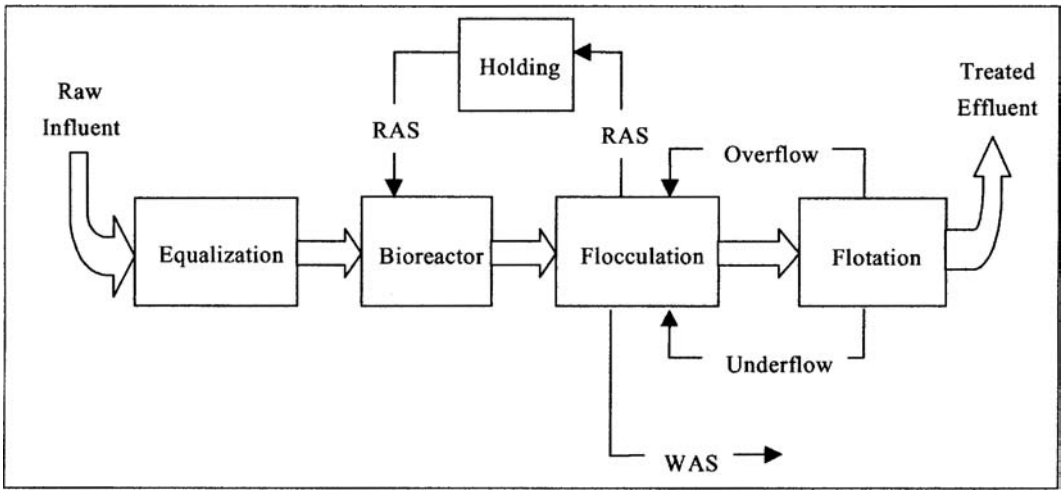

Fig. 2.15. Flow diagram of one train VERTREAT^{TM} VSB for the treatment of dairy wastewater (13)

the bioreactors in a recycle stream, and excess sludge is wasted to an off-site city dewatering facility. Treated effluent is discharged to the regional interceptor. Figure 2.15 is a simplified flow diagram that follows the wastewater flow through a single train of the plant. The second train of the plant is identical and there are multiple crossover points, providing complete redundancy.

9.1.2. Plant Loading and BOD Removal

Highly variable loading from the dairy wastewater averaged 45,000 lb BOD_5/d and the flow averaged 613 gpm. These values far exceeded the design capacity of the plant, which is rated for an average load and flow of 16,000 lb BOD_5/d and 486 gpm, respectively. Peak 4-h loading in the plant exceeded 25,000 lb BOD_5—more than five times the design value. Despite BOD_5 loading of 200% above design, the system achieved an average BOD_5 removal of 90% (11, 13). Influent and effluent wastewater composition and plant performance are shown in Table 2.4.

As shown in Table 2.6, the BOD_5 daily average, peak day, and peak 4-h period values were all far in excess of the plant design basis. While the design average and peak day values were exceeded by approximately 200%, the 4-h peak BOD_5 loading was exceeded by approximately 500%. These excessive loads necessitated additional aeration capacity in the plant. Additional aeration capacity enabled an average BOD_5 removal of 40,398 lb/d from a plant that was originally designed to remove 14,312 lb/d.

In addition to the extra aeration capacity, the excessive organic loading was offset by changes to the F/M ratio in the system. The shaft bioreactors were designed to operate with an MLSS of 7500 mg/L and an F/M ratio of 0.75/d. Operation at an MLSS of 7500 mg/L with the realized plant load would result in an average F/M ratio of 1.7/d, and a peak of 2.6/d. Operation at F/M ratios greater than 1.0/d can result in poor treatment, and can lead to foaming. In order

Table 2.4
Wastewater composition and performance of dairy VSB treatment plant (13)

Parameter	Unit	Design value	Actual value
Flow			
Daily average	gpm	486	613
Peak day	gpm	556	718
Peak 4-h period	gpm	556	975
Influent BOD_5			
Daily average	mg/L	2700	6170
Peak day	mg/L	3600	9220
Peak 4-h period	mg/L	3600	18,238
BOD_5 load			
Daily average	lb/d	16,000	45,000
Peak day	lb/d	24,032	68,274
Peak 4-h period	lb	4005	25,959
Temperature			
Daily average	°F	45 to 90	100
Peak day	°F	95	106
Peak 4-h period	°F	95	109
Influent pH	units	4 to 12	4.5 to 8.1
Effluent pH	units	6 to 9	7.6 to 8.3
BOD_5 removal			
Daily average	lb/d	14,312	40,398
Peak day	lb/d	22,363	59,958
Average BOD removal	%	90	90

to counteract the effects of the overload, the MLSS was raised to a value between 10,000 and 12,000 mg/L, effectively reducing the daily average F/M ratio to 1.1 to 1.3/d. These changes provided more buffering capability in the system for daily peak loads, and enabled operation with no persistent foaming events. A complication that arose from the increase in MLSS was the subsequent reduction in the A/S ratio (air to solids ratio) for flotation. These complications are discussed in detail in the following subsection on oxygen transfer and flotation.

As shown in Table 2.6, the average design flow for the facility was exceeded by 26%, and the peak 4-h period flow was 75% over design. This increase in the influent flow to the reactors resulted in a subsequent decrease in the HRT (hydraulic retention time), exacerbating the situation in the plant that was already dealing with an organic load 200% over design. The increased flow resulted in increased hydraulic and solids loading on the flotation clarifiers. The decreased retention time in the clarifiers, coupled with an increased solids flux rate, resulted in diminished flotation efficiency.

9.1.3. Oxygen Transfer Efficiency and Flotation

The VSB system achieves very high oxygen transfer efficiency (OTE)—greater than 75% OTE was measured routinely. These high oxygen transfer rates are associated with

the pressure and depth at which compressed air is introduced to the bioreactor. Despite an average oxygen transfer efficiency in excess of 70%, dissolved oxygen was limited in the system due to the overloaded conditions (the four installed 450-scfm compressors can provide approximately 32,000 lb O_2/d). The overloaded conditions in the plant necessitated additional aeration capacity. A diesel compressor with 1,000 scfm aeration capacity was installed as a temporary measure. This unit was eventually replaced with an electric compressor capable of delivering 1350 scfm. With the additional compressor, an average oxygen transfer efficiency of 73.4% was attained at 100°F, providing a total of approximately 48,000 lb O_2/d. While oxygen was still limited in the system improvements were noted with respect to flotation and dewatering (11, 13).

With the organic loading still well above design, and the aeration capacity operating at maximum, other measures were required to improve system stability and robustness. In order to minimize the effects of shock loading to the system, the MLSS concentration was increased in the bioreactors from the design value of 7,500 to 12,000 mg/L. The higher MLSS concentration decreased the F/M ratio and increased the buffering capacity of the system. While the decreased F/M ratio tended to improve microbe flocculation, this change also resulted in a corresponding 37% decrease in the A/S ratio (air to solids ratio), so no subsequent improvement in flotation performance was noted.

When operated at design loads, air, and dissolved oxygen are available in excess of the BOD_5 requirement in the system, providing the required air for proper flotation. During the overloaded conditions, all residual air in the reactor was utilized for BOD_5 destruction, leaving little or no dissolved oxygen for flotation. This also resulted in a decrease in the A/S ratio, creating a thinner float sludge blanket and reducing the solids capture efficiency in the clarifiers. Rapid increases in the organic load to the system and the corresponding F/M ratio resulted in dispersed bacteria. In the presence of proper amounts of BOD_5, bacteria tend to produce natural polymers that result in strong floc formations. In the presence of excess BOD_5, as encountered during operation, production of these polymers was likely lessened, resulting in a thinner float blanket. Deterioration of the flotation in the system necessitated polymer use to ensure that the mixed liquor solids concentration did not decrease below acceptable levels. If the MLSS were allowed to decrease in the system, the F/M ratio would increase, compounding the problem of the organic overload (11, 13).

9.1.4. Nutrient Limitation

The most important nutrients required for bacterial growth are phosphorus and nitrogen. Aerobic activated sludge systems require a minimum of approximately 1 mg of phosphorus and 5 mg of nitrogen for every 100 mg of BOD_5 removed from the wastewater. A deficiency in either the phosphorus or nitrogen supply can result in poor system operation. The effect of a nutrient limitation on a plant is usually severe. One or 2 h of insufficient nitrogen may result in upwards of 12 h of impaired BOD_5 removal, and up to 48 h of poor solids separation.

While the supply of phosphorus is more than adequate, dairy effluent is typically deficient in nitrogen content. Therefore, addition of an appropriate source of nitrogen is required prior to treatment. Piping and control logic were provided in the design for nutrient addition.

9.1.5. Temperature

The VSB system is capable of operating both in the mesophilic temperature range, 25°C to 38°C (77°F to 100°F), and in the thermophilic range, 46°C to 60°C (115°F to 140°F), while maintaining high oxygen transfer efficiency. There are advantages and disadvantages to both modes of operation.

Operation at mesophilic temperatures tends to foster a very robust, stable system that can handle loading, pH, and temperature swings that are considered intolerable in conventional activated sludge systems. The disadvantage of a mesophilic system is a relatively high sludge production of approximately 0.45 lb biomass per lb BOD_5 removed.

Advantages of thermophilic operations include no influent cooling requirements, lower sludge production, and decreased nutrient requirements. Disadvantages include less system stability because thermophilic organisms are sensitive to small variations in the environmental conditions, and diminished solids separation since the microbial population consists mainly of bacteria.

The VERTREAT™ plant at this dairy was designed to allow possible future conversion to thermophilic operation; materials were selected to allow operation at sustained temperature of 66°C (150°F). During the currant operation, the process was intended to operate in the mesophilic temperature range, 25°C to 38°C (77°F to 100°F). A BOD_5 of 2,700 mg/L (the design average influent BOD_5) will generate an approximate temperature rise of 7°C (12°F) in the bioreactor. The design influent temperature range was 38°C to 50°C (100°F to 122°F). Since mesophilic processes do not operate well above 35°C (95°F), influent coolers were added to cool the influent at an approximate rate of 3.5 MW (12 million BTU/h).

The anticipated bioreactor temperature rise due to BOD_5 degradation was 7°C (12°F). Due to the severe overloading in the plant, this temperature rise was closer to 22°C (40°F) on average, and reached as high as 28°C (50°F). Due to this excessive heat release, at times the heat exchangers were unable to keep bioreactor temperatures below 38°C (100°F). Operation at these temperatures fostered the growth of thermophilic microbes (mainly dispersed bacteria) that did not flocculate well and were more difficult to separate in the clarification stage. The ability of the system to achieve an average oxygen transfer efficiency of 73.4% at 38°C (100°F) was truly remarkable (11, 13).

9.1.6. VERTREAT™ Process Simplicity and Stability

Despite the overload on the system, it was found throughout the operation to be relatively simple to operate, resistant to upset, and able to rapidly recover from disruptions. This is attributed to enhanced oxygen transfer and pH buffering in the system, allowing treatment and neutralization of fluctuating waste loads.

During the periods of operation at design loads (typically a 3 to 4 h window during the day), air and dissolved oxygen were found to be available in excess of the BOD_5 requirement. This enabled operation of the bioreactor with 3 to 5 mg/L of dissolved oxygen. The availability of this additional dissolved oxygen facilitated the buffering of the extreme diurnal swings in waste strength observed at this plant.

During startup it was found that enough carbon dioxide was produced by the microbes—and subsequently held under pressure in the bioreactor—to neutralize intermittent pH levels as high as 12 (11, 13). This is well beyond the capability of any other biological treatment system. Conventional biological treatment systems require pH control in the range of 7 to 9. In a dairy plant such as this, with frequent caustic cleaning of equipment resulting in pH as high as 12 in the waste stream, pH control in that range would prove to be very costly. Although a pH controller and neutralization chemical pump were included in the design in case high influent pH levels were sustained for extended periods, they were not required during currant operation.

9.2. Refinery Wastewater Treatment

9.2.1. Plant Description

The major unit processes in the plant include secondary treatment using a vertical subsurface bioreactor, flotation clarifiers to separate the biological sludge from the treated effluent, effluent biofilters to remove residual suspended solids and refractory compounds, a vertical U-tube mesophilic aerobic digester placed within the reactor, and off-gas biofilters to treat all vent air from the process.

A flow diagram of the integrated Chevron refinery treatment plant is shown in Fig. 2.16. Influent to the bioreactor is supplied from a holding pond and contains sufficient nutrients for bio-oxidation in the reactor. The bioreactor provides the environment for a high rate activated biosolids system where the influent is mixed, aerated, contacted with return activated solids (RAS) and circulated. The reactor is operating in a temperature range close to the upper limit for mesophilic conditions, 27°C to 32°C (89°F to 90°F). To promote biodegradation and air lift circulation, air is injected into the downcomer and riser sections of the reactor (*see* Fig. 2.16). In addition to raw refinery effluent, biosolids (about 30% of normal influent flow) from the RAS holding tank are returned to the reactor. Return solids maintain an MLSS ranging from 2500 to 5000 mg/L in the reactor, which in turn supports an F/M ratio between 0.50 to 0.75 (11, 12, 32).

When the organic matter is stabilized, an extraction line approximately 76 m deep (250 ft) in the shaft is used to move material to the flocculation chambers ahead of the dissolved air flotation clarifiers. The dynamic head pressure from the head tank, vent stacks, and off-gas biofilters are calibrated to maintain optimum extraction velocities out of the reactor. Since dissolved air from depth is used for heterogeneous bubble nucleation in the flotation clarifiers, an optimal extraction velocity is controlled to prevent premature gas dissolution as the biodegraded material is extracted to ambient pressure regimes at the surface. In a simultaneous extraction from the head tank to the flocculation chamber, mixed liquor containing mostly dispersed gas bubbles completes the mixture by providing a balance between dispersed and dissolved gases prior to the dissolved air flotation clarifiers.

The dissolved air flotation clarifiers utilize three-phase separation of the incoming stream to provide recycle/waste solids, liquid effluent, and gases. The solids are either recycled to seed the bioreactor, or pumped to the aerobic digester where they are biodegraded to soluble organics and CO_2, and water. Thickened solids are wasted to the aerobic digester at a

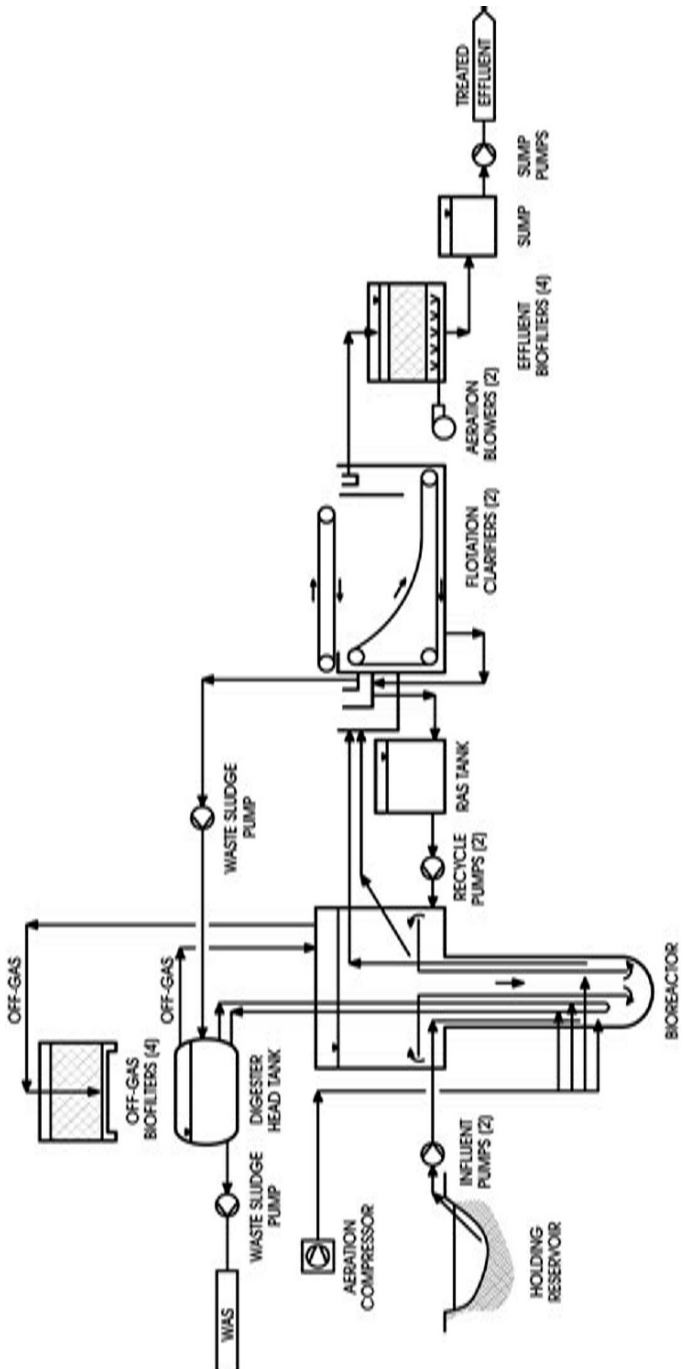

Fig. 2.16. Flow diagram of refinery wastewater treatment plant (12)

concentration ranging between 2.7% and 4.0% solids. Off the digester, a decant line returns subnatant to the holding pond or bioreactor and thickened biosolids are removed via vacuum truck to on-site thickening tanks. The digested solids are gravity thickened in the tanks to approximately 11% to 12% total solids prior to disposal in a dedicated landfill owned by the refinery.

The gas stream generated from the bioreactor and digester head tanks is fed to up-flow off-gas biofilters (biological aerated filters), which consist of attached growth on flooded porous media. These biofilters are designed to stabilize any compounds (such as VOCs) that are volatilized or otherwise escape treatment from the bioreactor and digester, ensuring that no odors are released from the treatment facility.

Clarified liquid effluent from the flotation clarifiers is sent to down-flow, fixed media effluent biofilters containing acclimatized biomass on porous media. The effluent biofilters polish any remaining organic material (refractory compounds) that has been slow to biodegrade. After biofiltration, the effluent flows to an effluent diversion tank, where it is either routed to a tank as filter backwash water, or is discharged to the municipal treatment works.

The primary function of the subsurface vertical shaft bioreactor is to remove organic compounds (BOD) from the refinery effluent. The shaft casing is 1.8 m (6 ft) in diameter and over 105 m (344 ft) deep, sealed at the bottom with a dished head. The entire unit is grouted into place. The reactor contains a cylindrical downcomer inside the main reactor. The air used for biological oxidation is injected at depth and provides a driving force for circulation at a controlled rate up the outer annulus and down the central downcomer. Effluent is injected at depth and is withdrawn below the injection point such that the effluent must make at least one circuit of the reactor before even a fraction is withdrawn. The flow transitions from up-flow to down-flow in a surface head tank at the top of the reactor.

The bioreactor head tank is directly connected to the riser and downcomer and utilizes a horizontal de-gas plate positioned such that the mixed liquor flows upward from the riser of the shaft, circulates to the end of the head tank under the de-gas plate, then returns on the top side of the plate and back to the downcomer. Gas that comes out of solution during this circulation is collected in the top of the head tank in one of the four compartments that are formed by longitudinal baffles on the roof of the head tank. Each of these compartments is connected directly to its own individual off-gas filter for treatment of VOCs and foam in the off-gas stream. The patented head tank and baffle systems apply the required hydraulic head pressure over the bioreactor to create the necessary extraction line velocities. Off-gas and foam from the digester tank and RAS from the flotation clarifiers are also directed to the bioreactor head tank.

The Chevron treatment plant has two flotation clarifiers to separate the biological sludge from the treated effluent. The two clarifiers measure $4.3 \times 15.8 \times 4.0$ m sidewall depth ($14 \times 52 \times 13$ ft). The patented flotation process is different from conventional methods in that, due to high pressure at depth in the reactor, the microbial mass (approximately 70% water) contains sufficient dissolved gas that, when the mass is brought to the surface from the deep part of the bioreactor, it is less dense than the surrounding liquid media. This assists the biosolids in flotation. A stream of mixed liquor extracted from the reactor will spontaneously separate into a thick float blanket and a clear liquid phase as dissolved gas

is released from solution. Heterogeneous bubble nucleation (dispersed gas acting as nuclei for dissolved gas attachment) is also an important phenomenon in this process. The floating biosolids thicken into a blanket with 30,000 to 40,000 mg/L total solids consistency (i.e. 3% to 4% solids, approximately two to three times the concentration of settled sludge in a conventional clarifier).

Clarified effluent is moved over a weir at the end of the flotation tank, and can go directly to the municipal treatment works (Greater Vancouver Regional District), to the refractory biofilters for polishing treatment, or to the backwash storage tank.

The majority of organics (approximately 80% to 90%) in petrochemical refinery effluent are easily biodegraded in the bioreactor. The remaining fraction biodegrades more slowly and is termed 'refractory' compounds. This fraction is most efficiently treated on attached growth biofilters that have the ability to attain a long sludge (matured or acclimatized) age of the biomass growing on the filter media. The principle mechanisms at work on the filter are (11, 12):

(a) Entrapment of the solid particles
(b) Sorption of colloidal material
(c) Bio-oxidation of the soluble organics

The first two mechanisms are physical-chemical and the efficiency can be improved with the use of chemicals such as alum or polymer. The third process, bio-oxidation, is accomplished by providing a favorable environment for the attached growth biomass.

Incorporated into this treatment plant is a 'U'-tube mesophilic aerobic digester set within the bioreactor to digest waste sludge. The digester consists of a 0.46 m (1.5 ft) downcomer and a 0.61 m (2 ft) riser connected at the bottom with a transitional 'U' bend, and at the top with the digester head tank. The digester 'U'-tube extends to a depth of 98.5 m (323 ft).

The digester consists of an aerated 'U'-tube with air-lift circulation and is suspended within the confines of the bioreactor casing. The upper ends of the tubes are in direct communication with the digester head tank, enabling circulatory flow. WAS enters the downcomer side of the 'U'- tube from the head tank. As in the bioreactor itself, the higher down-flow velocity will drag aeration bubbles in the downcomer to the bottom 'U' bend and return to the digester head tank with the coarser riser bubbles. The higher pressure in the bottom region tends to dissolve the air, thus providing the oxygen for the microbes in the stabilization process. Like in the bioreactor, the aeration in the digester aids in several other process functions at no incremental cost. It meets the microbial requirements for solids stabilization, provides circulation and mixing, and saturates the solids with entrained gas for digester head tank flotation.

The flotation thickening of the stabilized biosolids is performed in batches in the digester head tank. A circulation stall and reversal sequence is initiated to move, as quickly as possible, a volume of stabilized sludge containing high levels of dissolved gas from the bottom of the digester to the head tank. This is accomplished by reducing the riser air to levels less than the downcomer air, thus slowing and eventually stalling circulation. Upon stalling of the circulation, the voidage is greater in the downcomer than the riser and circulation reverses direction. Once the gas entrained fluid occupies the downcomer side, all air is turned off, leaving differential density to move the entrained material to the head tank. Flotation in the

head tank occurs as in a flotation clarifier, where the subnatant decant is gravity drained to the holding pond, and the stabilized thickened biosolids, 11% to 12%, are typically applied to a dedicated landfill owned by Chevron.

In order to provide a fully integrated treatment facility that treats the liquid, gas, and sludge streams, four up-flow fixed media off-gas biofilters were added to treat process off-gas from the bioreactor and digester. The off-gas biofilters measure $3 \times 3 \times 4$ m high ($10 \times 10 \times 13$ ft). Influent to the off-gas biofilters originates with the waste gas streams of the bioreactor and digester head tanks, containing spent air, foam, entrained biomass and water vapor. The filter is an up-flow design where pressurized off-gas from the bioreactor head tank filters through attached growth media. During the biomass oxidation in the bioreactor and the stabilization of biosolids in the digester, volatile absorbed organics, dissolved organics, and other metabolites of the process are released from the cell mass into the air or liquid streams. Although most of these VOCs are treated in the bioreactor, a portion remains that has not been stabilized (including those produced from biosolids digestion) and these are treated in the off-gas biofilters.

9.2.2. Treatment Plant Discharge Criteria

The discharge permit to the Greater Vancouver Regional District municipal treatment system has specified limits on the waste stream effluent from the Chevron refinery. Regulated in the discharge specification are flow, BOD_5, TSS, NH_3-N, FOG, and pH (Table 2.5). Due to sustained performance well below the discharge specifications, the permits were lowered in July 1998.

The refinery is targeting a constant effluent quality that will allow direct discharge into the receiving water (a BOD_5/TSS of 10/10 or better). The environmental mission statement for the refinery listed this objective as the primary mandate for the treatment plant. An application for a direct discharge has been set in motion based on confidence that the treatment plant can meet the direct discharge into the receiving water quality criteria.

9.2.3. BOD_5 and TSS Removal Efficiency

The plant has routinely achieved an effluent having a BOD/TSS of 15/15 or better (11, 12, 32). The average BOD_5 in the effluent for a 4-month study period in 2002 was 3.5 mg/L

Table 2.5
Chevron refinery wastewater treatment plant discharge specifications (12)

		Monthly averages (mg/L)					
	Flow	BOD_5	TSS	NH_3	FOG	pH range	Toxicity
August 1997	2592 m^3/day[a] (.685 US MGD)	300[b]	100	10	10	6.0–10.5	N/a
July 1998	2592 m^3/day[a]	100[b]	40	10	10	6.0–10.5	$LC_{50} = 100\%$

[a] Maximum instantaneous discharge flow rate.
[b] Maximum concentration.

with a maximum BOD_5 of 9.4 mg/L. BOD_5 concentrations in the influent varied widely and hit a maximum of 295 mg/L during the period. This represented an average BOD_5 removal of 98% in the treatment system. The effluent was well below the discharge specification of 100 mg/L BOD_5 and was even below the direct discharge into receiving water of 10 mg/L.

The average TSS in the plant effluent for the same study period was 14.3 mg/L with a maximum TSS of 36.0 mg/L. This represented excellent solids removal in the flotation clarifier and effluent biofilters. TSS concentrations in the influent averaged 164 mg/L and hit a maximum of 716 mg/L during the period. This represented an average TSS removal of 86% in the treatment system. The effluent solids were well below the discharge specification of 40 mg/L

9.2.4. Solids Reduction Efficiency in the Aerobic Digester

Digester studies (August 2000) showed that there was approximately a 26% reduction in total solids and a 30% reduction of volatile solids across the digester. This is an exceptional amount of volatile solids destruction in the system considering the mesophilic digester operated at an average temperature of approximately 30°C, and the average solids detention time in the digester was only 4.5 days.

The enhanced level of destruction at shortened detention times is primarily attributable to the increased level of oxygen transfer in the digester. The amount of air reaching the microbes and circulating the sludge has a significant effect on the digester solids reduction efficiency. At present, the aeration rate is approximately 43 scfm, which means that an average oxygen transfer efficiency of approximately 35% is being achieved in the digester. It is certainly possible that the rate of aeration could be optimized for even further digestion.

9.2.5. pH Buffering

In the Chevron treatment system it has been noted that enough carbon dioxide is produced by the microbes—and subsequently held under pressure in the bioreactor—to neutralize pH levels ranging from 10 to 11.5 for extended durations, and intermittent pH levels as high as 12. The resulting bicarbonates provided a natural buffer within the mixed liquor and stabilized the operation against pH swings. The result was an effluent that is consistently buffered at a pH of 8.0.

This is well beyond the capability of any other biological treatment system (11, 12). Conventional biological treatment systems require strict pH control in the range of 7 to 9 to avoid system upset, reduced treatment efficiency, or an outright kill of the microorganisms. In a refinery plant such as this, there is relatively high use of caustic, primarily for the washing of hydrocarbon streams (i.e. for the extraction of H_2S from light hydrocarbon streams), and for the neutralization of acids used in the process. Any upsets of caustic washers can result in as high as a pH of 12 in the waste stream. pH control in that range could prove to be very costly in a conventional process. No pH controllers or neutralization chemical pumps were installed at the Chevron facility.

9.2.6. Removal of Toxicity and Recalcitrant Compounds

In order to meet direct discharge standards the effluent from the treatment plant has to pass fish toxicity testing (LC_{50} testing where the $LC_{50} = 100\%$); i.e. it must be non-toxic at full

strength (undiluted). Regular fish toxicity testing indicated that the effluent had consistently met this specification except for incidents of gross overloading in the plant. While there is a substantial amount of data that supports the conclusion that the overall treatment plant meets discharge requirements, the treatment efficiency of individual unit operations, such as the tertiary biofilters, was not originally quantified. In recent studies the operational performance of the Chevron refinery effluent biofilters has been tested in order to quantify the role these units play in acute toxicity removal.

The treatment system has the ability to consistently remove over 90% of the toxicity present in the refinery effluent stream (measured as EC_{50}). It should be noted that a Microtox result of 35% (measured as $EC_{50} = 35\%$) is equivalent to a Fish Toxicity result of 100% (measured as $LC_{50} = 100\%$). As expected, the majority of the toxicity (between 80% and 90%) was removed upstream of the effluent biofilters in the bioreactor and flotation cell. A further 2% to 10% of the overall reduction occurred through the biofiltration stage of the process. These results indicate that the biofilters are behaving as a 'polishing' stage, effectively reducing wastewater toxicity as an integral part of the overall treatment facility.

Ongoing testing has focused on quantifying the maximum capacity of the plant to remove toxic compounds from the wastewater. Tests to date have shown that the process is not only capable of removing persistent compounds such as phenols to very low concentrations (0.02 mg/L), but that compounds such as MTBE, which were thought to be virtually non-biodegradable, are partially degraded in the system.

9.2.7. Process Simplicity and Stability

Since commissioning in 1996, the performance of the system has been excellent. All discharge specifications were met easily when the plant was operating under normal conditions. The plant has also proved robust against swings in influent flow, strength, and pH, considerably outside the design basis. In particular, the buffering capacity of the reactor has proved truly remarkable and large caustic spills have been absorbed by the effluent treatment system with no action from plant personnel.

Despite occasional overloads on the system due to process upsets in the refinery, the treatment system has been relatively simple to operate, resistant to upset, and was able to rapidly recover from disruptions. This is attributed to enhanced oxygen transfer and pH buffering in the system, allowing treatment and neutralization of fluctuating waste loads (11, 12, 32).

9.3. Municipal Wastewater Treatment

The City of Homer's wastewater treatment plant was funded, in part, by a grant from the US Environmental Protection Agency. The funding was granted after EPA's evaluation of the technology and its approval to consider the Vertical Shaft Bioreactor system, with flotation clarification, as an innovative process based on energy savings and advancement of the state-of the-art (33–35).

Homer, Alaska wastewater treatment plant is the first municipal plant in the US to adopt the VSB technology. The plant has been in operation since 1991, and won the 1993 AWWA Large Plant of the Year award for the State of Alaska. The plant has met or exceeded specification since commissioning, and upon start-up it successfully passed a 1-year performance certifi-

cation program for the US EPA (36, 37). During this certification period, the plant achieved an average effluent quality 33% below discharge specifications. The average annual effluent BOD and TSS concentrations were 20 and 19 mg/L respectively.

9.3.1. Plant Description

Raw wastewater entering the treatment plant is pumped from the on-site pumping station to a rotary type, mechanically cleaned bar screen in the plant headworks. Screenings removed on the bar screen are dewatered in a screenings press and incinerated in a dedicated screenings incinerator.

From the bar screen, the wastewater flows to the grit chamber, which is of the cyclonic type with no moving parts. Grit removed in the unit can be dewatered and incinerated with screenings or pumped to a sludge lagoon. Effluent from the grit chamber is piped to a flow splitting structure that distributes the flow to the secondary process. Primary clarification is not provided (36). Figure 2.17 shows the plant process flow diagram.

Secondary treatment is provided by the Vertical Shaft Bioreactor process. The VSB process uses two vertical shafts, 2.5 ft (0.76 m) in diameter by 500 ft (152 m) deep, as aeration tanks. Each shaft is fitted with a 1.5 ft diameter concentric pipe (downcomer) that conveys the liquid to the bottom of the shafts. The liquid returns to the surface in the annular space surrounding the downcomer. Compressed air, injected into both areas of each shaft at the 200-ft depth, supplies oxygen to the process and provides circulation in the shafts through air-lift action. Because of the high pressure in the shafts, the oxygen transfer efficiency is very high.

The high pressure in the shafts also solubilizes the air. When the mixed liquor returns to the surface, the dissolved gas is released that causes solids to float. Flotation clarifiers are therefore used in place of gravity clarifiers for final solids separation. Polymers must be added to the mixed liquor to achieve efficient clarification. Effluent from the flotation clarifiers is disinfected using ultraviolet radiation prior to discharge to Kachemak Bay.

Waste activated sludge, removed by skimming the flotation clarifiers, is pumped to two aerobic digestion tanks. The digesters are aerated using a coarse bubble diffused air system. Digested sludge is discharged, by displacement, to a sludge lagoon. The lagoon is aerated using floating mechanical aerators of the aspirating propeller type design.

Sludge is removed from the lagoon during spring and summer using a floating dredge that pumps the sludge to covered sludge drying beds. Four of the fourteen drying bed cells are designed for composting dewatered sludge using the aerated static pile process.

9.3.2. Plant Design Criteria

Design criteria for the plant and BOD_5/TSS effluent requirements are presented in Table 2.6.

9.3.3. Plant Assessment

The overall performance of the VSB process during the US EPA mandated performance evaluation period was very good. The NPDES effluent solids concentration was exceeded only on the few occasions when abnormal conditions occurred. The process was temporarily affected by shock loadings, particularly from septage, but recovered very quickly (37).

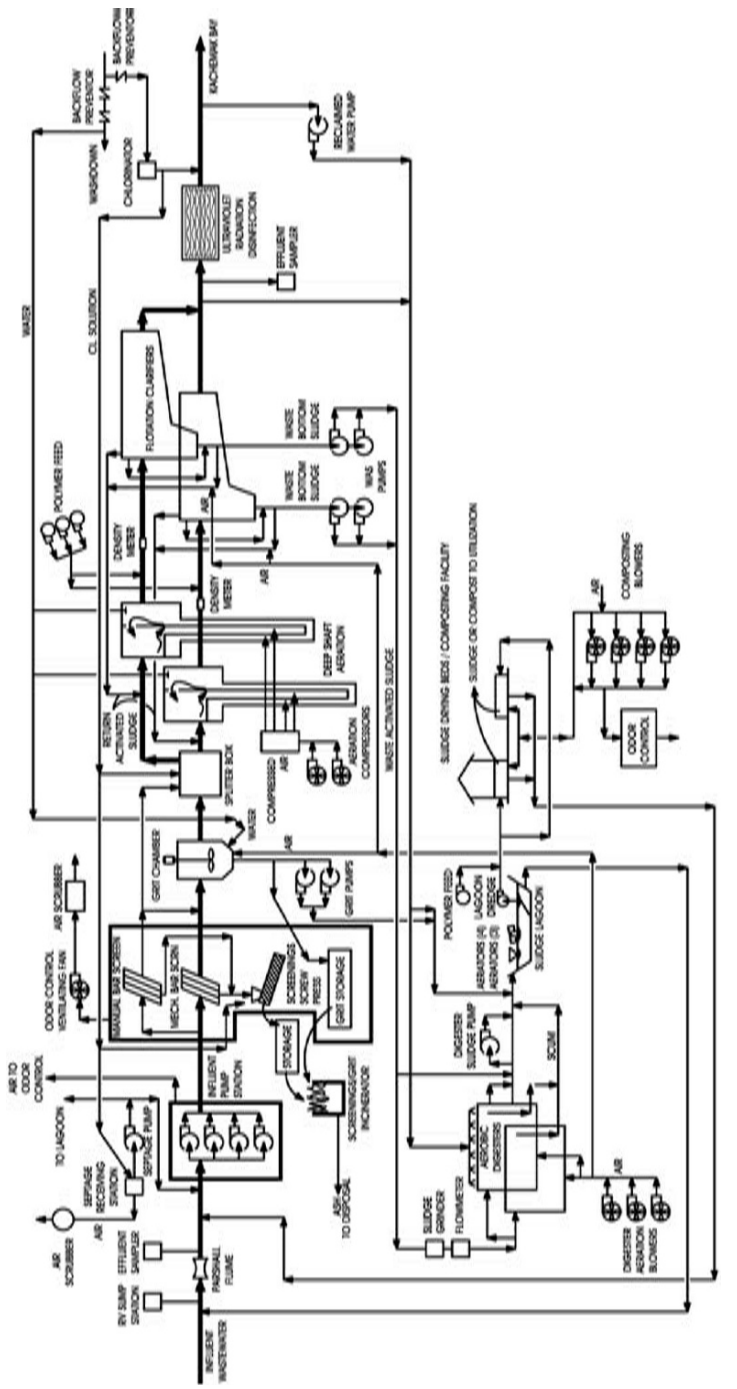

Fig. 2.17. Flow diagram of Homer City VSB wastewater treatment plant (41)

Table 2.6
Design, wastewater composition and performance of Homer VSB wastewater treatment plant (Extracted from Homer City WWTP Performance Certification Report 1992, Ref. 41)

Parameter	Unit	Design value	Actual value
Flow			
Daily average (peak month)	MGD	0.88	0.46
Peak day	MGD	1.43	0.90
Influent BOD_5			
Daily average (peak month)	mg/L	225	232
Peak day	lb/d	1645	749
Influent TSS			
Daily average (peak month)	mg/L	317	229
Peak day	lb/d	2320	754
Effluent BOD_5			
Daily average (peak month)	mg/L (lb/d)	30 (155)	20
Peak day	mg/L	60	—
Effluent TSS	mg/L (lb/d)	30 (155)	19
Daily average (peak month)			
Peak day	mg/L	60	—
Hydraulic detention time (t)	h	1.0	2.3
Mean cell residence time (MCRT)	day	2.0	1.6
Food to microorganism (F/M)	lb BOD/lb MLVSS	1.0	0.64
MLSS	mg/L	6600	4642
MLVSS	mg/L	5300	3749
Air requirement (flow/shaft)	scfm	130	—
Discharge pressure	psi	100	—
BOD removal	%	85	91
TSS removal	%	85	92

The plant was not operated near design loadings during the 1 year certification period. The highest monthly BOD_5 loading during the period was 60% of the design BOD_5 loading. Because the plant was operated at low MLSS concentrations the average monthly F/M ratio was equal to the design F/M ratio of 1.0 lb BOD_5/lb MLVSS/d. Furthermore, the MCRT was only 1.2 days, which is significantly less than the design MCRT of 2.0 days. In spite of these operating conditions, the average monthly effluent concentrations were well below permit requirements. These operating results indicate that the plant can handle significantly higher loadings without violating the discharge permit (36).

In order to treat significantly higher BOD_5 loadings, the plant must be operated at higher mixed liquor concentrations. During the monitoring period, the average MLSS concentration was 4642 mg/L. The design MLSS concentration is 6600 mg/L.

The aeration system functioned well throughout the period. The dissolved oxygen concentration in the mixed liquor was consistently very high and was typically above 15 mg/L. The lowest DO concentrations occurred in a summer month when the average concentration

dropped to 11.9 mg/L. Only one of the two aeration compressors was operated. Operations staff reported that aeration controls were simple and effective.

Based on the first year's operating results and experience, it had become obvious that proper operation of the flotation clarifiers was the key to good process performance. The most important and difficult aspect of clarifier operation was sludge wasting (38–40). It was so because sludge wasting had been a manual operation that required considerable operator attention. Modifications to the control system were made to allow programmed, automatic wasting. Operations staff found that this greatly improved the overall solids management of the VSB process and improved the solids concentration of the float.

Actual solids production was as predicted during design and about 10% less than typical for high rate activated sludge plant operating without primary clarification. Solids production would be reduced if the process was operated at higher MCRT.

Energy consumption for the VSB process has been very close to the consumption estimated during facilities planning. The power usage for the VSB was 1007 kWh/d compared to the 1549 kWh/d for the alternative conventional extended aeration process. This is an energy reduction of 29%, which is greater than the 20% required for the US EPA designation as an innovative technology (36).

Seasonal process impacts due to weather were less significant than expected. The temperature of the influent, which ranged from a monthly average low of 6.6°C in March to a high of 13.4°C in July, did not have a significant impact on process performance. Operations personnel, however, reported that solids separation was more difficult during low temperatures. One seasonal condition, which is likely to occur every year, was the heavy algae growth in the sludge lagoon during the long daylight hours of summer. Algae, which enter the influent to the plant via the return flow from the lagoon, were not removed by the treatment process and caused an increase in effluent suspended solids concentrations. Chlorination of the return flow is necessary to control the impact on the process (36).

In 2001 the Homer, Alaska plant had fully completed its 10th year in operation. Recent operating data from the plant that covers a full year of operations, between July 2000 and June 2001 (41) were obtained and are summarized in Table 2.7. The data shows that even the

Table 2.7
Wastewater composition and performance of Homer VSB treatment plant-10 years later (Extracted from Homer City WWTP operating data 2000 to 2001, Ref 46)

Parameter	Average	% of design value	Maximum
Flow (MGD)	0.58	65	1.13
BOD_5			
Influent (mg/L)	307	73	372
Effluent (mg/L)	13	43	19
Removal (%)	95	112	95
TSS			
Influent (mg/L)	360	88	663
Effluent (mg/L)	18	60	31
Removal (%)	95	112	95

maximum effluent BOD_5 for the whole period was just 19 mg/L, indicates that the plant must still be working fine 10 years into its operation.

Another thing to note is that the maximum flow through the plant was 0.58 MGD, which is 65% of the design capacity. This figure is misleading, however, because the Homer plant only operates with just one of the two installed clarifiers. This means that the single clarifier is being operated at 113% of its design capacity, and has done so more or less continuously since the plant was commissioned in 1991 (41).

For more information on the application of VSB to municipal wastewater treatment, the reader is referred to the Vertical Shaft-Flotation Plant for the City of Bangor, Maine (42, 43).

NOMENCLATURE

C = the minimum dissolved oxygen concentration, mg/L
C_{SW} = the oxygen saturation concentration in wastewater, mg/L
dc/dt = the rate of change in dissolved oxygen concentration, kg/m^3/h (lb/ft^3/h)
F = the driving force required to maintain flow through the reactor
F/M = food to microorganisms ratio = kg BOD_5/kg MLSS (lb BOD/lb MLSS)
H = depth of VSB, m (ft)
K_{LA} = the oxygen transfer rate coefficient, h^{-1}
MLSS = mixed liquor suspended solids, mg/L
MLVSS = mixed liquor volatile suspended solids, mg/L
Q_i = influent waste flow rate, m^3/h (ft^3/h)
Q_R = mixed liquor flow rate through the VSB, m^3/h (ft^3/h)
t = mean residence in the reactor, h
TSS = total suspended solids, mg/L
v = flow-through velocity inside the reactor, m/h (ft/h)
V_G = volume of gas bubbles, m^3 (ft^3)
V_L = volume of liquid, m^3 (ft^3)

REFERENCES

1. S. M. Shere and P. G. Daly, Deep Shaft treatment of thermo-mechanical pulping effluents, *Canadian Pulp and Paper Association, Environment Improvement Conference*, Quebec City, Quebec, Canada, 21–23 October (1980).
2. D. S. Sanford and J. J. Faria, *Application of Deep Shaft Technology to Food Processing Wastes*, Eco-Research Ltd., Toronto, Ontario, Canada (1980).
3. D. S. Sanford and D. A. Chisholm, The treatment of municipal wastewaters using the ICI Deep Shaft process, *Proceedings of the 29th Annual Western Canada Water and Sewage Treatment Conference*, Edmonton, Alberta, Canada (1977).
4. US EPA, *Technology Assessment of the Deep Shaft Biological Reactor*, Report # EPA-600/2-82-002, US Environmental Protection Agency, Municipal Environmental Research Laboratory, February (1982).

5. E. H. Dunlop, Characteristics of sludge produced by the deep shaft process, Reprint from the Society of Chemical Industry, *New Processes of Wastewater Treatment and Recovery Conference*, Belgrave Square, London (1977).
6. H. M. Bailey, J. C. Ousby, and F. C. Roesler, ICI Deep Shaft aeration process for effluent treatment, *Proceedings of the Institution of Chemical Engineers symposium*, Series No. 41 (1975).
7. B. J. Bradley, ICI Deep Shaft effluent treatment process, Proceedings of the 29th Annual Atlantic Canada Section, ANWA Meeting, Hosting the 6th Annual PACE Meeting, Halifax, Nova Scotia, Canada (1976).
8. R. C. Brenner and J. J. Convery, Status of Deep Shaft wastewater treatment technology in North America, *Proceedings, 7th United States/Japan Conference on Sewage Treatment Technology*, Tokyo, Japan (1980).
9. D. H. Bolton, D. A. Hines, and J.P. Bouchard, The application of the ICI Deep Shaft process to industrial effluents, *Proceedings of the 31st Annual Purdue Industrial Waste Conference*, Lafayette, Indiana (1976).
10. R. I. Dick, Folklore in the design of final settling tanks, *Journal of the Water Pollution Control Federation*, 48, 4, April (1976).
11. NORAM Bio Systems, *Information Package-Videos, Flyers, Papers, Presentations CD*, NORAM Engineering and Construction Ltd., Vancouver, BC, Canada, March (2003)
12. J. Guild and D. Pollock, Refinery wastewater treatment utilizing the VERTREAT™ Process, *WEF Industrial* (2003).
13. J. Guild and D. Pollock, Startup of a full-scale VERTREAT™ wastewater treatment plant at a cheese production facility, *WEFTEC* (2001).
14. Metcalf and Eddy, *Wastewater Engineering Treatment and Reuse*, 4th ed., McGraw-Hill, New York (2003).
15. W. W. Eckenfelder, *Principles of Water Quality Management*, CBI Publishing Company, Boston, MA (1980).
16. Editor, New Units Groomed for Sewage, Garbage, *Canadian Chemical processing*, 5 February (1979).
17. B. Appleton, ICI's Deep Shaft halves sludge volume, *New Civil Engineer*, 17 April (1975).
18. P. G. Daly and C. C. Shen, The deep shaft biological treatment process, *43rd Purdue Industrial Waste Conference Proceedings*, Lewis Publishers, Chelsea, MI (1989).
19. D. Pollock, *Long Vertical Shaft Bioreactors for Biosolids Treatment*, Text transmitted by e-mail from NORAM Engineering and Construction Ltd., Vancouver, BC, Canada, April (2004).
20. Eco-Research, Ltd., *Ithaca Deep Shaft, Visitor's Program*, Eco-Reseatch, Ltd., Willowdale, Ontario, Canada (1980).
21. R. E. Speece, Energy Efficiency of Deep U-Tube Aeration, Proceedings of the Department of Energy, Energy Optimization of Water and Wastewater Management for Municipal and Industrial Applications Conference, Argonne National Laboratory, Argonne, IL (1979).
22. B. Appleton, Thurrock, Test-bed for ICI Deep Shaft, *New Civil Engineer*, 14 April (1977).
23. ICI, The ICI Deep-Shaft Effluent Treatment Process, *Engineers' Digest*, 36, 5, May (1975).
24. The Deep-shaft process: bright future for treating high-strength wastewaters, *Journal of Civil Engineering-ASCE*, 51, 5 May (1981).
25. J. P. Bouchard, F. B. Knudsen, S. Lessard, and L. W. Keith, North American application of the Deep Shaft effluent treatment process, *Proceedings of the 50th Annual Conference, WPCF*, Philadelphia, Pennsylvania, 2–7 October (1977).

26. WEF-Buyer's Guide, *Deep Shaft Technology Inc. Equipment*, Calgary, AB Canada, Web Site Address: www.wef.org/applications/buyersguide/company.cfm?CompanyID=845 (2004).
27. W. L. Wardrop and Associates, *Comparison of Extended Aeration and Deep Shaft Sewage Treatment Plants*, City of Portage La Prairie, Prepared by W. L. Wardrop & Associates Ltd., Engineering Consultants, Canada, June (1977).
28. City of Portage La Prairie, *Sewage Treatment*, Submission to The Clean Environment Commission, Canada, 19 June (1978).
29. G. E. Thompson, Selection of Deep Shaft technology for the City of Portage La Prairie, Manitoba, *Proceedings of the Workshop 79, New Developments in Wastewater Treatment*, University of Toronto, Toronto, Canada, 7–8 March (1979).
30. US EPA, Areawide Assessment Procedures Manual, Point Source Control Alternatives: Performance and Cost, US Environmental Protection Agency, MERL, Cincinnati, OH, July (1976).
31. US ACE, *Civil Works Construction Cost Index System Manual*, 110-2-1304, US Army Corps of Engineers, Washington, DC, PP 44, http://www.nww.usace.army.mil/cost, (2008).
32. J. Ng, R. Freeman, and D. Pollock, Refinery wastewater treatment utilizing an innovative Deep Shaft Process, *53rd Annual Purdue University Industrial Waste Conference*, West Lafayette, Indiana (1998).
33. US EPA, *Technical Review of Deep Shaft Innovative Designation Request for Homer, Alaska*, US Environmental Protection Agency, Office of Research and Development, Water Engineering Research Laboratory, Cincinnati, OH (1987)
34. US EPA, *Innovative and Alternative Technology Assessment Manual*, EPA-430/9-78-009, US Environmental Protection Agency, Cincinnati, OH (1980).
35. J. M. Smith, J. J. McCarthy, and H. L. *Longest, Impact of Innovative and Alternative Technology in the United States*, Municipal Environmental Research Laboratory, US Environmental Protection Agency, Cincinnati, OH (1980).
36. R. W. Beck and Associates, Inc., *One Year Performance Certification Program-City of Homer WWTP*, US EPA Grant # C-020133-01, Final Report, September (1992).
37. NORAM Bio Systems, *VERTREAT & VERTAD Endorsement CD*, NORAM Engineering and Construction Ltd., Vancouver, BC, Canada, January (2003).
38. W. J. Brignal and G. C. Cox, The deep shaft method of sewage treatment, *Cabedeau International Conference*, Liege, Belgium, May (1979).
39. R. C. Brenner, Status of novel biological process development in the United States, Proceedings of the Interim Technical Seminar Between Seventh and Eighth US/Japan Conferences on Sewage Treatment Technology, Tokyo, Japan, 15 May (1981).
40. Eco-Research, Ltd., *Commissioning and Operation of Deep Shaft Demonstration Plant, Ithaca, New York*, Prepared by: Eco-Research Limited, Willowdale, Ontario, Canada (1980).
41. Homer WWTP, *Homer, Alaska WWTP Operating Data for the Period July 2000 to June 2001*, Data sheet provided by Dave Pollock, NORAM Engineering and Construction Ltd. April (2004).
42. L. K. Wang and P. G. Daly, *Preliminary Design Report of a 10-MGD Deep Shaft-Flotation Plant for the City of Bangor, Maine*, Technical Reports No. PB88-200597/AS, US Department of Commerce, National Technical Information Service, Springfield, VA (1987).
43. L. K. Wang and P. G. Daly, *Preliminary Design Report of a 10-MGD Deep Shaft-Flotation Plant for the City of Bangor, Maine: Appendixes*, Technical Reports No. PB88-200605/AS, US Department of Commerce, National Technical Information Service, Springfield, VA (1987).

APPENDIX

United States Yearly Average Cost Index for Utilities US Army Corps of Engineers (31)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

3
Aerobic Granulation Technology

Joo-Hwa Tay, Yu Liu, Stephen Tiong-Lee Tay, and Yung-Tse Hung

CONTENTS

Abstract Recently, attention has been given to aerobic granulation, which is a novel environmental biotechnology for wastewater treatment. This chapter reviews the progress and development of basic research and application of aerobic granular sludge sequencing batch reactors in the treatment of a wide variety of wastewaters.

Key Words Aerobic granulation • sequencing batch reactor • wastewater treatment.

1. INTRODUCTION

Microbial granulation is a process of cell-to-cell self-immobilization involving biological, physical, and chemical actions. Granules formed through self-immobilization of the microorganisms are dense consortia packed with different bacterial species that typically contain millions of organisms per gram biomass. These bacteria perform different roles in degrading the complex industrial wastes containing various organic chemicals, nitrogen, and phosphorus. Compared to the conventional activated sludge, the granules have a regular, dense, and strong microbial structure, good settling property, high biomass retention, and ability to withstand high-strength wastewater and shock loading.

Granulation occurs in both aerobic and anaerobic wastewater treatment systems. Formation of anaerobic granules has been studied for decades, and is probably best recognized in the

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_3

upflow anaerobic sludge blanket (UASB) reactor. Hundreds of plants worldwide currently employ the anaerobic granulation technology (1, 2). However, application of this technology is greatly limited by drawbacks such as the long start-up period required (normally 2 to 8 months), a relatively high operation temperature, and unsuitability for low-strength organic wastewater (2). To overcome these weaknesses, recent research efforts have been dedicated to developing aerobic granulation technology for the removal of organic wastes. The development of aerobic granules was first reported by Mishima and Nakamura (7) in a continuous aerobic upflow sludge blanket reactor. Aerobic granules with diameters of 2 to 8 mm were developed, with good settling properties. Aerobic granulation has since been reported in sequencing batch reactors (SBRs) by many researchers (3–6, 8–11), and can be applied in high-strength organic wastewater treatment, carbon, and nitrogen removal as well as toxic wastewater treatment (12, 13). This chapter reviews key findings concerning the aerobic granulation technology, and describes the current state of knowledge about the aerobic granulation process, the structure and microbial diversity of aerobic granules, and the suitability of aerobic granulation for various wastewater treatment applications.

2. AEROBIC GRANULATION AS A GRADUAL PROCESS

The formation of aerobic granules in SBR was tracked by using advanced image analysis techniques, and was shown to be a gradual process. Dispersed seed sludge with a mean size of about 100 μm developed into small aggregates, which evolved into compact granular sludge, then finally matured into aerobic granules with a mean size >0.25 mm (Figs. 3.1 and 3.2). The seed sludge exhibited a typical morphology of conventional activated sludge, with a very loose, fluffy, and irregular structure in which filamentous organisms were present (Figs. 3.1A and 3.1B). Compact and dense sludge aggregates appeared after 1 week of reactor operation (Fig. 3.2), while granular sludge with clear round outer shapes formed after 2 weeks of reactor operation (Fig. 3.2). In week 3, mature aerobic granules dominated the whole reactor

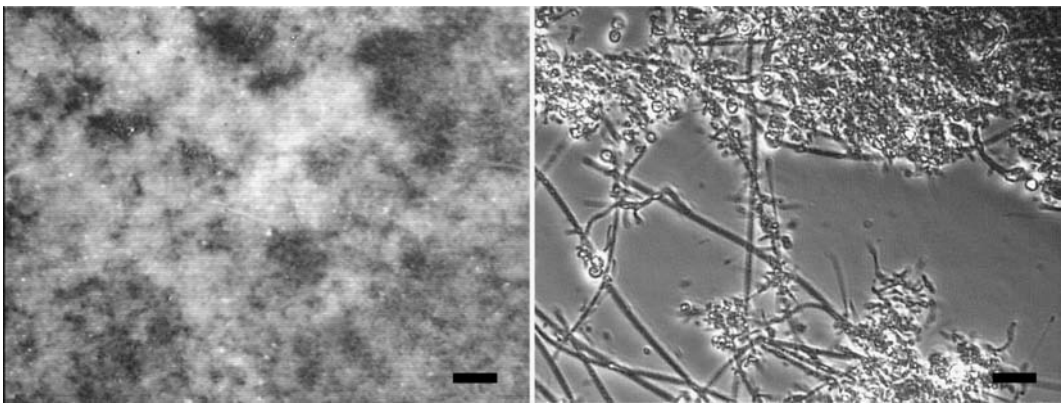

Fig. 3.1. Morphology of seed activated sludge used for cultivation of aerobic granules. A: viewed by image analysis. *Bar*: 2 mm; B: viewed by optical microscope. *Bar*: 5 μm. (*Source*: Adapted from (5)).

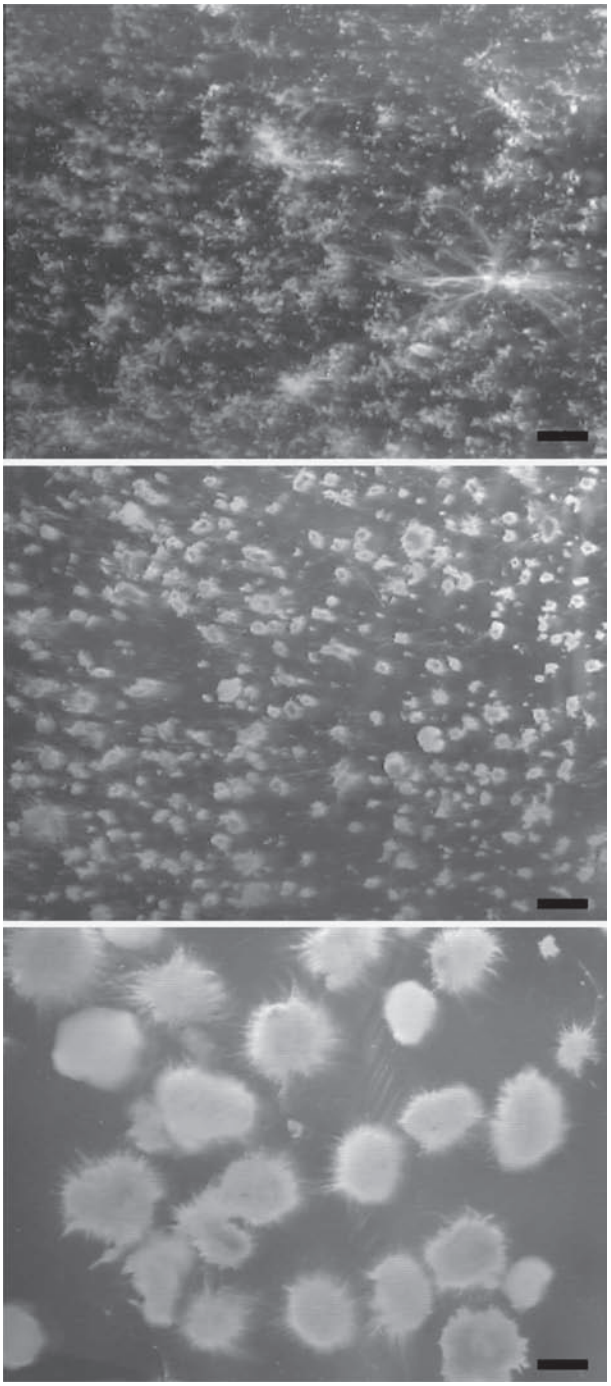

Fig. 3.2. Image analysis of the sludge morphology at different operation time. *Bar*: 2 mm. (*Source*: Adapted from (5)).

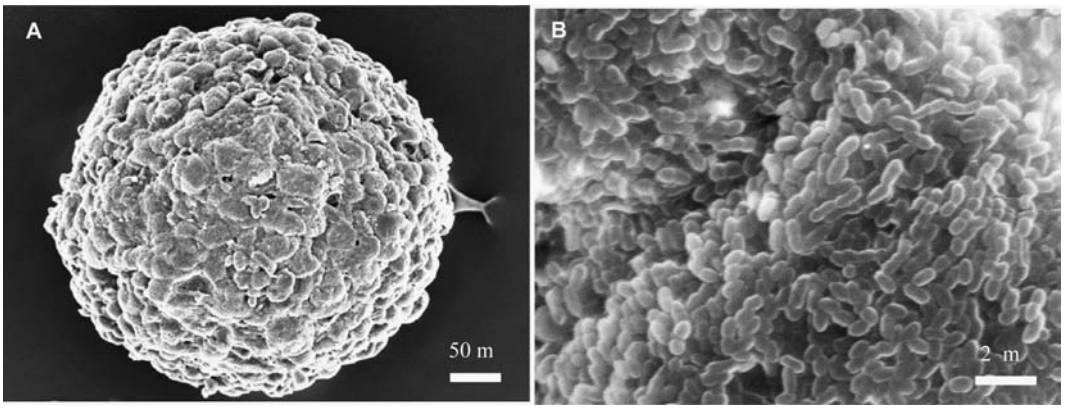

Fig. 3.3. Scanning electron micrograph of aerobic granule (A) and its surface microstructure (B).

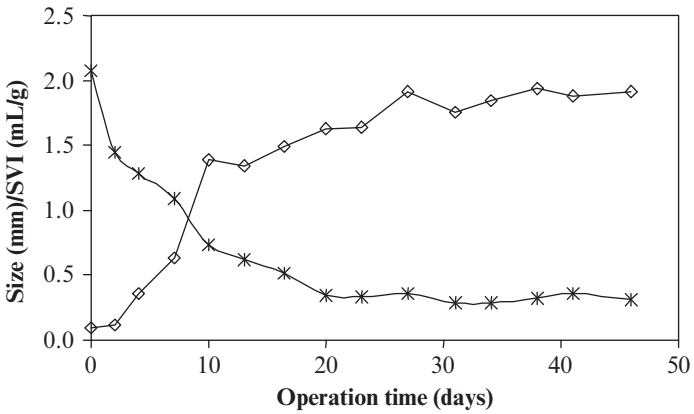

Fig. 3.4. Sludge size (◊) and SVI (∗)(×100) versus the operation time in the course of aerobic granules cultivation.

(Fig. 3.2), and had a very regular round-shaped outer structure. Scanning electron micrograph (SEM) images of aerobic granules grown on acetate as sole carbon source revealed a compact microbial structure in which individual cells were tightly linked up together (Fig. 3.3). Sludge volume index (SVI) measurements showed that mature aerobic granules possessed significantly improved sludge settleability compared to the initial seed sludge (Fig. 3.4).

3. FACTORS AFFECTING AEROBIC GRANULATION

3.1. Substrate Composition

The essential role of carbon source in the formation of anaerobic granules has been demonstrated (14, 15). In the case of aerobic granulation, experimental evidence suggests that aerobic granulation seems to be insensitive to the nature of substrate carbon source; for

example, aerobic granules had been successfully cultivated with a wide variety of substrates, including glucose, acetate, ethanol, phenol, and synthetic wastewater (5, 8, 12, 16). However, granule microstructure and species diversity appears to depend on the type of carbon source. The glucose-fed aerobic granules exhibited a filamentous structure (Fig. 3.2), whereas acetate-fed aerobic granules had a nonfilamentous and very compact bacterial structure, in which a rodlike species was predominant (Fig. 3.3). It should be pointed out that aerobic granules could also be cultivated with nitrifying bacteria and an inorganic carbon source (17). These nitrifying aerobic granules showed excellent nitrification ability.

3.2. Organic Loading Rate

The organic loading rate (OLR) is one of the most important parameters in the design and operation of biological wastewater treatment facilities. The essential role of organic loading rate in the formation of anaerobic granules has been widely recognized. A relatively high organic loading rate facilitated the formation of anaerobic granules in UASB systems (18–21). In contrast to anaerobic granulation, the accumulated evidence suggests that aerobic granules can form across a wide range of organic loading rates, from 2.5 to 15 kg COD/m^3 day, i.e., aerobic granulation is less dependent upon the organic loading rate applied (3, 13, 16). This is probably attributable to the nature of aerobic bacteria.

Although the effect of organic loading rate on the formation of aerobic granules is insignificant, the physical characteristics of aerobic granules are dependent on organic loading rate. The mean size of aerobic granules increased from 1.6 to 1.9 mm with the increase of the organic loading organic loading from 3 to 9 kg COD/m^3 day (16). This is simply attributable to the fast growth of aerobic bacteria at high organic loading rates. A similar trend was also observed in anaerobic granulation (2, 22). It seems that the growth patterns of both aerobic and anaerobic granules under different organic loading rates are subject to the classical Monod model. The effect of organic loading rate on the morphology of mature aerobic granules in terms of roundness was found to be insignificant, whereas the aerobic granules developed at different organic loading rates exhibited comparable dry biomass density, specific gravity, and SVI. On the other hand, the physical strength of aerobic granules decreased with the increase of organic loading rate (16). Similarly, in anaerobic granulation process, it was also found that a high organic loading rate resulted in a reduced strength of anaerobic granules, i.e., partial loss of structural integrity and disintegration would occur at high organic loading rate (23, 24). In fact, an increased organic loading rate may raise the biomass growth rate, and high growth rate of microorganisms in turn would reduce the strength of the three-dimensional microbial community structure. Consequently, organic loading rate plays an important role in maintaining the stability of aerobic granules.

3.3. Hydrodynamic Shear Force

The contribution of hydrodynamic shear to anaerobic granulation in UASB had been reported (1, 25–27), and its essential role in biofilm process has attracted intense research attention (28–30). A high shear force results in biofilms with a strong and compact microbial structure, whereas a weak shear force produces biofilms with a heterogeneous and porous structure (28–32). Shear force also plays a very important role in the formation of aerobic

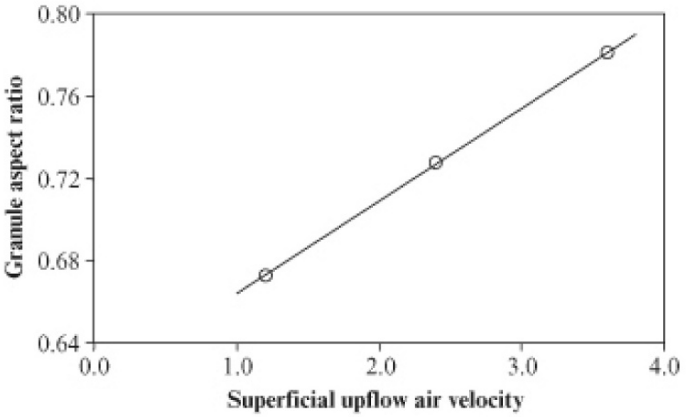

Fig. 3.5. Effect of superficial upflow air velocity on granule morphology.

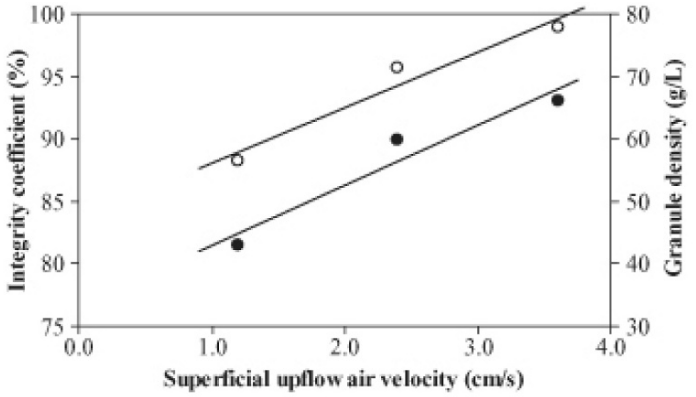

Fig. 3.6. Effect of superficial upflow air velocity on granule density (•) and physical strength in terms of integrity coefficient (○).

granules. A high shear force favors the formation of aerobic granules and granule stability (33, 34). It was found that aerobic granules could be formed only above a threshold shear force value in terms of superficial upflow air velocity above 1.2 cm/s in a column SBR, and more regular, rounder, and compact aerobic granules were developed at high hydrodynamic shear force (16, 34). Fig. 3.5 shows that the aspect ratio of acetate-fed aerobic granules increased with shear force in terms of superficial upflow air velocity, i.e., granules would become rounder when shear force increased, whereas the granule density and strength that represent the compactness of a microbial community were also proportionally related to the shear force applied (Fig. 3.6). These may imply that the structure of aerobic granules is mainly determined by the hydrodynamic shear force present in the bioreactor. In fact, the effect of shear force on granule structure is similar to its effect on biofilm, i.e., higher shear force leads to a thinner and denser biofilm (28, 30, 32, 35, 36).

It is well-known that extracellular polysaccharides can mediate both cohesion and adhesion of cells and play a crucial role in maintaining structural integrity in a community of immobilized cells (29, 37–39). Tay et al. (39) reported that the production of extracellular polysaccharides (PS) was closely associated with the shear force. The extracellular polysaccharide content normalized to proteins (PN) increased with the shear force in terms of superficial upflow air velocity, i.e., high shear force stimulated bacteria to secrete more extracellular polysaccharides (Fig. 3.7). In fact, shear force-induced production of extracellular polysaccharides had been commonly observed in biofilm process (40–42). Consequently, the enhanced production of extracellular polysaccharides at high shear can contribute to the compact and stronger structure of aerobic granules (34, 39). The metabolic network of cells includes interrelated catabolic and anabolic reactions. The catabolic activity of microorganisms is directly correlated with the electron transport system activity, which can be described by the SOUR. Tay et al. (34) reported that the SOUR of aerobic granules increased with the increase of shear force (Fig. 3.8). It is most likely that the shear force can stimulate

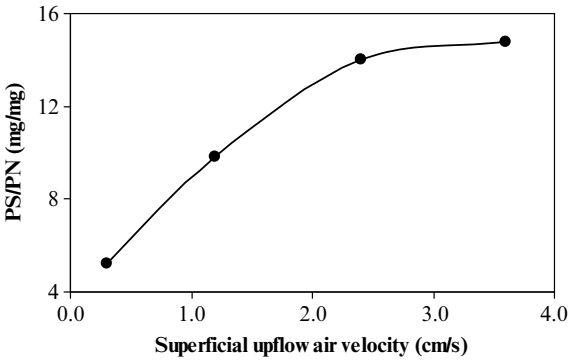

Fig. 3.7. Effect of superficial upflow air velocity on the production of cell polysaccharides (PS) normalized to cell proteins (PN). (*Source*: Adapted From (34)).

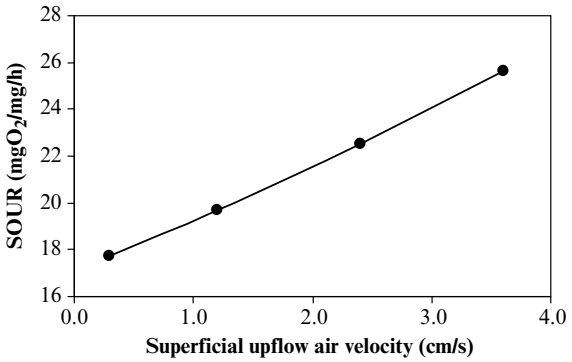

Fig. 3.8. Effect of superficial upflow air velocity on sludge specific oxygen uptake rate (SOUR). (*Source*: Adapted From (34)).

microbial respiration activity. It appears from Figs. 3.7 and 3.8 that when the shear force is increased, much more energy generated by catabolism would be used for the production of extracellular polysaccharides rather than for microbial growth. This in turn indicates that when shear force exerted on granular sludge is high, the granules would have to regulate their metabolic pathway so as to maintain a balance with the external shear force, through consuming nongrowth-associated energy, i.e., the microbial community may respond to shear force by metabolic changes and some biological events might be involved in shear-associated phenomena (16, 28).

3.4. Presence of Calcium Ion in Feed

Polyvalent cations such as Ca^{2+}, Mg^{2+}, and Fe^{2+} have been suggested for stimulating the anaerobic granulation process by neutralizing negative charges on bacterial cell surfaces, thus creating relatively strong "van der Waals" attractive forces (43–45). Ca^{2+}, with concentration of 100 to 200 mg/L, was found to exert a positive impact on anaerobic granulation (46, 47). Grotenhuis et al. (48) found that granules disintegrated or became weaker after treatment with a Ca^{2+} chelating agent, EGTA (ethylene glycol-bis (β-aminoethyl ether)-*N*, *N*,-tetraacetic acid). These seem to imply that calcium ions might play an important role in anaerobic granule structure as calcium phosphate precipitates for adhesion of bacteria. Similar observations were reported for aerobic granulation. Jiang et al. (49) found that the addition of Ca^{2+} accelerated the aerobic granulation process. With the addition of 100 mg Ca^{2+}/L, the formation of aerobic granules took 16 days compared to 32 days in the culture without the Ca^{2+} addition. The Ca^{2+} augmented aerobic granules also showed better settling and strength characteristics, and had higher polysaccharide content. It had been proposed that that Ca^{2+} could bind to negatively charged groups present on bacterial surfaces and extracellular polysaccharide molecules, and act as a bridge to interconnect these components and promote bacterial aggregation. Polysaccharides play an important role in maintaining the structural integrity of biofilms and microbial aggregates such as aerobic granules, as they are known to form a strong and sticky nondeformable polymeric gellike matrix, and can contribute to cell-to-cell adhesion through interactions between secondary functional groups such as hydroxyl and calcium ions.

3.5. Reactor Configuration

Reactor configuration will have an impact on the flow pattern of liquid and microbial aggregates in the reactor (3, 29). Column-type upflow reactors and completely mixed tank reactors (CMTR) have very different hydrodynamic behaviors in terms of interactive patterns between flow and microbial aggregates. The air or liquid upflow pattern in column reactors can create a relatively homogenous circular flow along the reactor height, and microbial aggregates are constantly subject to such a circular hydraulic attrition. The circular flow could force microbial aggregates to be shaped as regular granules that have a minimum surface free energy. In a column-type upflow reactor, a higher ratio of reactor height to diameter (H/D) can ensure a longer circular flowing trajectory, which in turn creates a more effective hydraulic attrition to microbial aggregates (3, 29). However, in CMTR, microbial aggregates stochastically move with dispersed flow in all directions. Thus, microbial aggregates are subject to varying localized hydrodynamic shear force, flowing trajectory and random collision.

Under such circumstances, only flocs of irregular shape and size instead of regular granules occasionally form (29). Therefore, the column-type reactor with high ratio of reactor height to diameter, which can provide an optimal interactive pattern between flow and microbial aggregates, is favorable for the formation of aerobic granules.

3.6. Dissolved Oxygen

Dissolved oxygen (DO) concentration is an important parameter in the operation of aerobic wastewater treatment systems. Aerobic granules formed at the DO concentration as low as 0.7 to 1.0 mg/L in a SBR (4), whereas they were also successfully developed at high DO concentrations up to 5 mg/L. It appears that DO concentration would not be a decisive parameter in the formation of aerobic granules.

4. MICROBIAL STRUCTURE AND DIVERSITY

4.1. Characteristics of Aerobic Granule

The physical characteristics and microbial activity of aerobic granules are summarized in Table 3.1.

Morphology: Compared to conventional bioflocs, aerobic granules have a defined spatial shape. The average roundness in terms of aspect ratio is higher than 0.6 for aerobic granules grown on different carbon sources. As discussed earlier, the roundness of aerobic granules is mainly influenced by the external shear force. The mean diameter of mature aerobic granules varies, and depends on the substrate composition, organic loading rate, shear force, etc.

Settleability: The settling property of aerobic granules is a key operation factor that determines the efficiency of solid–liquid separation, and it is essential for the proper functioning of wastewater treatment systems. Both SVI and settling velocity are used to describe the sludge settleability. The SVI of aerobic granules is much lower than that of conventional bioflocs (Table 3.1). This implies that, from an engineering perspective, the settleability of sludge can be improved significantly through the formation of aerobic granules, and a more compact clarifier would be adequate. The settling velocity of aerobic granules is associated with granule size and structure. The settling velocity of aerobic granules is usually higher than 30 m/h, which is comparable with that of the UASB granules, and is at least three times higher than that of activated sludge flocs, which have a typical settling velocity of around 8 to 10 m/h.

Granule density and strength: The specific gravity of aerobic granules falls into a range of 1.004 to 1.065. The granules with high physical strength would have a strong ability to withstand high abrasion and shear. The physical strength, expressed as integrity coefficient (%), which is defined as the ratio of residual granules to the total weight of the granular sludge after 5 minutes of shaking at 200 rpm on a platform shaker, is higher than 95% for the aerobic granules grown on glucose and acetate. This indicates that the physical strengths of aerobic and anaerobic granules are comparable.

Cell surface hydrophobicity: The cell surface hydrophobicity was 68% for glucose-fed aerobic granules and 73% for acetate-fed granules. These values are two times higher than that

Table 3.1
Characteristics of aerobic granules

Substrate	Average diameter (mm)	Average aspect ratio	SVI	Settling velocity (m/h)	Integrity coefficient (%)	Cell surface hydrophobicity (%)	Biomass density (g/L)	Specific gravity (kg/L)	SOUR ($mg\ O_2/g\ h$)	Reactor type	References
Sewage	2.0–8.0		40.8–143							AUSB[a]	(7)
Glucose, acetic and yeast	0.5–2.5			86.4						AUSB[a]	(33)
Molasses	2.35			30–40					96.5	SBR	(8)
Sodium acetate	0.3–0.5		80–100							SBR	(4)
	0.5–1.6		20–45					1.007	76.2	SBR	(73)
Ethanol	1.9–4.6	0.69–0.86					11.9			SBR	(3)
Sodium acetate	1.0	0.7								SBR	(10)
Glucose, peptone, acetate	3.0			72				1.04–1.054		SBR	(11)
Glucose	2.4	0.79	51–85	35	97	68	41.1		69.4	SBR	(52)
Acetate	1.1	0.73	50–80	30	98	73	32.2		55.9	SBR	(52)
Phenol	0.52		40						110	SBR	(12)

[a] Aerobic upf.

of the seed sludge. There is strong evidence showing that cell surface hydrophobicity plays a crucial role in cell-to-surface attachment and cell-to-cell self-immobilization (16, 29, 50–52).

Granule storage stability: Similar to anaerobic granules, aerobic granules exhibited good storage stability. Aerobic granules cultivated with glucose showed a 60% reduction in microbial activity after 4 months of storage at 4°C (52). The loss in microbial activity of aerobic granules by storage would be associated with the length of storage time, the type of feed carbon, and the culture history. Granules maintained a good shape and showed little reduction in physical strength in terms of integrity coefficient after 4 months of storage. Zhu and Wilderer (53) also found that after 7 weeks of storage of aerobic granules in ambient environment, the aerobic granules could regain microbial activity within a week.

4.2. Layered Structure of Aerobic Granules

Confocal laser scanning microscopy (CLSM) was used with different oligonucleotide probes, specific fluorochromes, and fluorescent microspheres to study the microstructure of aerobic granules (54–57). The obligate aerobic ammonium-oxidizing bacteria *Nitrosomonas* spp. was found mainly at a depth of 70 to 100 μm from the granule surface, and aerobic granules contained channels and pores that penetrated to a depth of 900 μm below the granule surface. The porosity in granules peaked at depths of 300 to 500 μm from the granule surface. These channels and pores would facilitate the transport of oxygen and nutrients into and metabolites out of the granules. Polysaccharide formation peaked at a depth of 400 μm below the granule surface. The anaerobic bacteria *Bacteroides* spp. was also detected at a depth of 800 to 900 μm from the granule surface, whereas a layer of dead microbial cells was located at a depth of 800 to 1000 μm. To fully use the aerobic microorganisms in the granules, the optimal diameter for aerobic granules should be less than 1,600 μm, which is twice the distance from the granule surface to the anaerobic layer (54). Consequently, smaller granules will be more effective for aerobic wastewater treatment as these granules have more live cells within a given volume of granules.

4.3. Microbial Diversity of Aerobic Granules

The microbial diversity of aerobic granules is closely related to the composition of culture media in which they are developed. Glucose-fed aerobic granules mainly consisted of filaments and some cocci bacteria, whereas rod-shaped bacteria were dominant in granules grown on acetate (5, 52). By using ribosomal-based molecular techniques and PCR-cloning, Yi et al. (58) detected shifts in microbial diversity among young, mature, and old aerobic granules cultivated on glucose. The development of aerobic granules appeared to be a dynamic process that involved an assemblage of microorganisms. Shifts in microbial diversity were attributed to physiological adaptation by various microorganisms during the aerobic granulation process. Changes in bacterial composition and species abundance would be attributed to interactions among different groups of bacteria and the microniches that they occupy. Microorganisms associated with five operational taxonomic units were found in all granules sampled at different stages of development, which suggests that these bacteria may play an important role in the development of aerobic granules. Different operational taxonomic units were also

found to dominate different growth stages. For example, several types of microorganisms were dominant in the old granules, and not detected in the young granules. This finding is important as changes in relative abundance may be used as markers of granule development, or even reflect the onset of granule lysis and deterioration.

5. MECHANISM OF AEROBIC GRANULATION

For bacteria to form granules, a number of conditions must be met, and the contributions of physical, chemical, and biological forces to the granulation process should be considered jointly. Liu and Tay (29) proposed a generalized model for the granulation process as follows:

Step 1: Physical movement to initiate bacterium-to-bacterium contact. The forces involved in this step are:

- Hydrodynamic force.
- Diffusion force.
- Gravity force.
- Thermodynamic forces, e.g., Brownian movement.
- Cell mobility. Cells can move by means of flagella, cilia, and pseudopods, whereas cell movement may also be directed by a signaling mechanism.

Step 2: Initial attractive forces to keep stable multicellular contacts. Those attractive forces are:

Physical forces:

- Van der Waals forces
- Opposite charge attraction
- Thermodynamic forces including free energy of surface; surface tension
- Hydrophobicity
- Filamentous bacteria that can serve bridge to link or grasp individual cells together

Chemical forces:

- Hydrogen liaison
- Formation of ionic pairs
- Formation of ionic triplet
- Interparticulate bridge and so on

Biochemical forces:

- Cellular surface dehydration
- Cellular membrane fusion
- Signaling and collective action in bacterial community

Step 3: Microbial forces to make cell aggregation mature:

- Production of extracellular polymer by bacteria, such as exopolysaccharides, etc.
- Growth of cellular cluster
- Metabolic change and genetic competence induced by environment, which facilitate the cell–cell interaction, and results in a highly organized microbial structure

Step 4: Steady state three-dimensional structure of microbial aggregate shaped by hydrodynamic shear forces.

The microbial aggregates would be finally shaped by hydrodynamic shear force to form a certain structured community. The outer shape and size of microbial aggregates are determined by the interactive strength/pattern between aggregates and of hydrodynamic shear force, microbial species and substrate loading rate, etc.

It should be emphasized that the hydrophobicity of bacterial surface plays a potentially important role in the initiation of aerobic granulation. According to thermodynamic theory, increasing the hydrophobicity of cell surfaces would cause a corresponding decrease in the excess Gibbs energy of the surface, which in turn would promote cell-to-cell interaction and further serve as a driving force for bacteria to selfaggregate out of the liquid phase (hydrophilic phase). It has been pointed out that hydrophobic binding has a prime importance for cell attachment (50, 59). A high hydrophobicity of the cell surface would result in a stronger cell-to-cell interaction and a denser structure (76). It has been generally believed that cell surface hydrophobicity is essential to the formation of biofilms and anaerobic granules (47, 51, 60). It was reported that cell surface hydrophobicity of aerobic granules was much higher than that of sludge flocs (5, 52, 61). Therefore, cell surface hydrophobicity might also play an important role in aerobic granulation.

Cell polysaccharides can mediate both cohesion and adhesion of cells and play a crucial role in maintaining the structural integrity in a community of immobilized cells, such as biofilms and anaerobic granules (38, 62, 63). In fact, in the study of anaerobic granulation, Harada et al. (64) observed that the extracellular polymers excreted by acidogenic bacteria appeared to enhance the strength and structural stability of anaerobic granules. Vandevivere and Kirchman (65) also found that the content of exopolysaccharides was five times greater for attached cells than for free-living cells. It was reported that colanic acid, an exopolysaccharide of *Escherichia coli* K-12, is critical for the formation of the complex three-dimensional structure and depth of *E. coli* biofilms (66). The polysaccharide contents of aerobic granules are much higher than that of sludge flocs (34, 39). Cell polysaccharides would also contribute greatly to aerobic granulation.

6. APPLICATIONS OF AEROBIC GRANULATION TECHNOLOGY

6.1. High-Strength Organic Wastewater Treatment

Granulation of the sludge could lead to high biomass retention in the reactor because of the compact microbial structure of granules. Biomass concentrations as high as 6.0 to 12.0 g/L have been obtained in SBRs operating with a volumetric exchange ratio of 50% (6, 52). The feasibility of applying aerobic granulation technology for the treatment of high-strength organic wastewater was demonstrated by Moy et al. and Tay et al. (13, 75, 76), who examined the ability of aerobic granules to sustain high organic loading rates by introducing step increases in organic loading only after COD removal efficiencies have stabilized at values greater than 89% for at least 2 weeks. In this way, aerobic granules cultivated on glucose were exposed to organic loading rates that were gradually raised from 6.0 to 9.0, 12.0, and 15.0 kg COD/m^3 day. Aerobic granules were able to sustain the maximum organic loading rate of 15.0 kg COD/m^3 day employed, and attained COD removal efficiencies greater than 92%. The granules initially exhibited a fluffy loose morphology dominated by filamentous

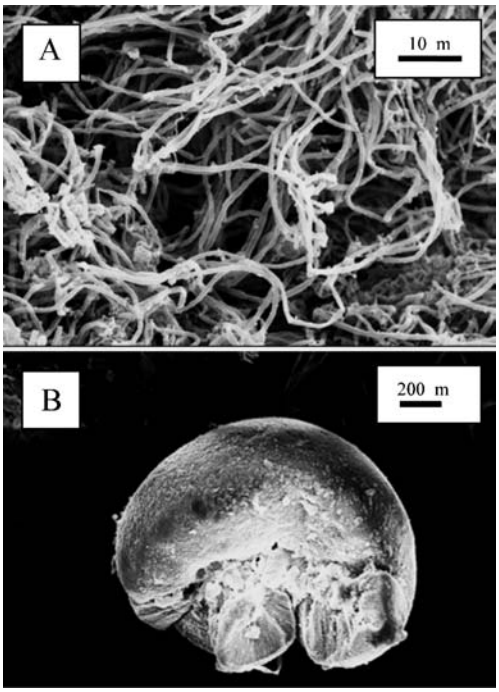

Fig. 3.9. Scanning electron micrographs of glucose-fed granules. A: 6 Kg COD/M^3 D and 750× magnification. B: 15 Kg COD/M^3 D and 37× magnification.

bacteria at low loading, and evolved into smooth irregular shapes characterized by folds, crevices, and depressions at higher loading (Fig. 3.9). These irregularities were thought to allow for better diffusion and penetration of nutrients into the granule interior. Diffusion was also enhanced by the higher substrate concentration that existed in the bulk solution at higher loading. These factors enabled the aerobic granules to sustain high organic loading rates without compromising granule integrity.

6.2. Phenolic Wastewater Treatment

Phenol is a commonly employed chemical in many industries. Because of its widespread use, however, phenol is a major pollutant in many industrial wastewaters, and its removal from wastewater is a subject of obvious importance. Phenol-containing wastewater is difficult to treat because of substrate inhibition. Microbial growth on phenol substrate and concomitant phenol biodegradation are hindered by the toxicity exerted by high concentrations of the substrate itself. However, the selfimmobilization or aggregation of microbial cells into compact granules could serve as an effective protection against high phenol concentrations. Jiang et al. were first to demonstrate that aerobically grown microbial granules could be successfully cultivated to degrade phenol (12). These phenol-degrading aerobic granules have excellent phenol biodegradation ability. For an influent phenol concentration of 500 mg/L, a stable

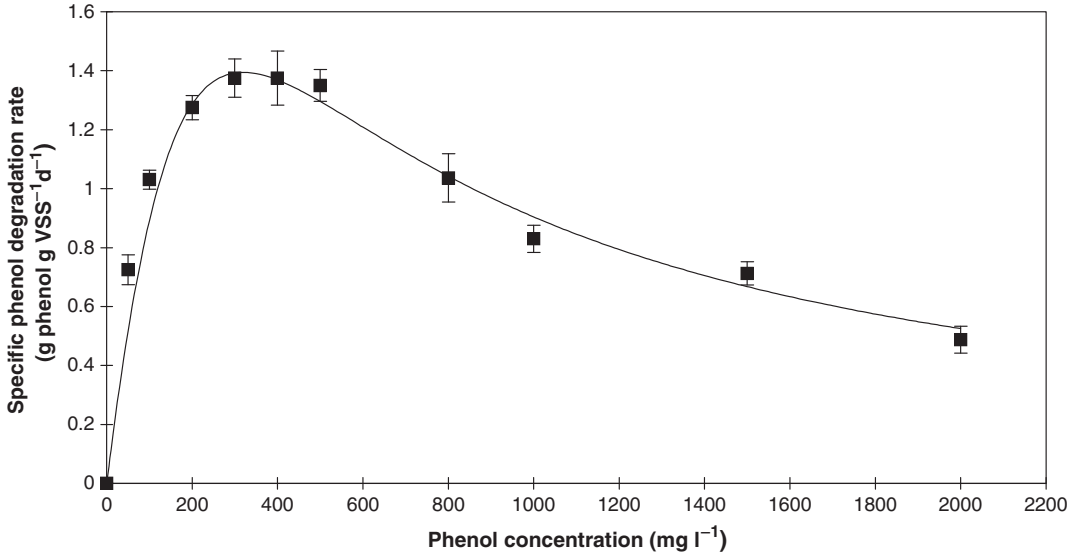

Fig. 3.10. Specific phenol degradation rates of microbial granules at different phenol concentrations.

effluent phenol concentration of less than 0.2 mg/L was achieved in the aerobic granular sludge reactor (12). The phenol-degrading aerobic granules had a specific phenol degradation rate as high as 1.4 g phenol/g MLVSS day, which was two times higher than that of acclimated seed sludge (Fig. 3.10). The kinetic behavior of the phenol-degrading granules followed the well-known Haldane model (12), indicating that the phenol-degrading aerobic granules developed a phenol uptake system that counteracted the adverse effects of phenol inhibition. Although the specific phenol degradation rates peaked at 500 mg phenol/L and gradually declined thereafter, significant rates of phenol degradation were still attained at phenol concentrations as high as 2,000 mg/L. This high tolerance of aerobic granules for phenol can be exploited to develop compact high-rate aerobic granulation systems for the treatment of industrial wastewaters containing high phenol concentrations. It can be expected that aerobic granules would be powerful bioagents for the removal of inhibitory or toxic organic compounds from high-strength industrial wastewaters.

6.3. Biosorption of Heavy Metals by Aerobic Granules

Heavy metals are often found in a wide variety of industrial wastewaters. More stringent metal concentration limits are being established because of the relatively high toxicity of heavy metals to environmental receptors. A vast array of biomaterials have been tested as biosorbents for heavy metal removal, such as marine algae, fungi, hairy roots of *Thlaspi caerulescens*, waste-activated sludge, digested sludge and so on (67–70). In view of the physical characteristics of aerobic granules as discussed earlier, aerobic granules are ideal for removing heavy metals in wastewater because of their strong microbial structure with large surface area and high porosity. Moreover, aerobic granules would be easily separated from the

liquid after biosorption capacity is exhausted. The biosorption of heavy metals such as zinc(II) and cadmium(II) by aerobic granules has been shown by Liu et al. (71, 72). It was found that the biosorption of Zn(II) by aerobic granules was related to both initial Zn(II) and granule concentrations (71), with the concentration gradient of Zn(II) as a main driving force for Zn(II) biosorption on the surfaces of aerobic granules. The biosorption capacity of aerobic granules would be related to the ratio of initial Zn(II) concentration to the initial granule concentration. It was found that the maximum biosorption capacity of Zn(II) by aerobic granules is 270 mg/g, whereas for Cd(II) biosorption, the maximum capacity is 566 mg/g (71, 72). Consequently, the aerobic granule-based biosorption process is an efficient and cost-effective technology for the removal of heavy metals from industrial wastewater.

NOMENCLATURE

CLSM = confocal laser scanning microscope
CMTR = completely mixed tank reactor
COD = chemical oxygen demand, mg/L
DO = dissolved oxygen, mg/L
EGTA = ethylene glycol-bis (b-aminoethyl ether)-*N*, *N*,-tetraacetic acid
H/D = reactor height to diameter ratio
MLVSS = mixed liquor volatile suspended solids, mg/L
PN = proteins, mg/L
PS = extracellular polysaccharides, mg/L
SBR = sequencing batch reactors
SEM = scanning electron microscopy
SOUR = specific oxygen uptake rate, mg O_2/g h
SVI = sludge volume index, mL/g
UASB = upflow anaerobic sludge blanket reactor

REFERENCES

1. M. Alves, A. J. Cavaleiro, E. C. Ferreira, A. L. Amaral, M. Mota, M. da Motta, H. Vivier, and M. N. Pons, Characterization by image analysis of anaerobic sludge under shock conditions, *Water Science and Technology*, 41, 207–214 (2000).
2. Y. Liu, H. L. Xu, K. Y. Show, and J. H. Tay, Anaerobic granulation technology for wastewater treatment, *World Journal of Microbiology and Biotechnology*, 18, 99–113 (2002).
3. J. J. Beun, A. Hendriksm, M. C. M. van Loosdrecht, E. Morgenroth, P. A. Wilderer, and J. J. Heijnen, Aerobic granulation in a sequencing batch reactor, *Water Research*, 33, 2283–2290 (1999).
4. D. Peng, N. Bernet, J. P. Delgenes, and R. Moletta, Aerobic granular sludge-a case study, *Water Research*, 33, 890–893 (1999).
5. J. H. Tay, Q. S. Liu, and Y. Liu, Microscopic observation of aerobic granulation in sequential aerobic sludge blanket reactor, *Journal of Applied Microbiology*, 91, 168–175 (2001).
6. J. H. Tay, Q. S. Liu, and Y. Liu, Aerobic granulation in sequential sludge blanket reactor, *Water Science and Technology*, 46, 13–18 (2002).

7. K. Mishima and M. Nakamura, Self-immobilization of aerobic activated sludge – a pilot study of the aerobic upflow sludge blanket process in municipal sewage treatment, *Water Science and Technology*, 23, 981–990 (1991).
8. E. Morgenroth, T. Sherden, M. C. M. van Loosdrecht, J. J. Heijnen, and P. A. Wilderer, Aerobic granule sludge in a sequencing batch reactor, *Water Research*, 31, 3191–3194 (1997).
9. J. J. Beun, M. C. M. van Loosdrecht, and J. J. Heijnen, Aerobic granulation in a sequencing batch airlift reactor, *Water Research*, 36, 702–712 (2002).
10. J. J. Beun, M. C. M. van Loosdrecht, E. Morgenroth, and J. J. Heijnen, Aerobic granulation, *Water Science and Technology*, 41, 41–48 (2000).
11. E. Etterer and P. A. Wilderer, Generation and properties of aerobic granular sludge, *Water Science and Technology*, 43, 19–26 (2001).
12. H. L. Jiang, J. H. Tay, and S. T. L. Tay, Aggregation of immobilized activated sludge cells into aerobically grown microbial granules for the aerobic biodegradation of phenol, *Letters in Applied Microbiology*, 35, 439–445 (2002).
13. B. Y. P. Moy, J. H. Tay, S. K. Toh, Y. Liu, and S. T. L. Tay, High organic loading influences the physical characteristics of aerobic sludge granules, *Letters in Applied Microbiology*, 34, 407–412 (2002).
14. W. M. Wu, *Technological and microbiological aspects of anaerobic granules*. Ph.D. thesis, Michigan State University, MI, USA (1991).
15. J. Chen and S. Y. Lun, Study on mechanism of anaerobic sludge granulation in UASB reactors, *Water Science and Technology*, 28, 171–178 (1993).
16. J. H. Tay, Q. S. Liu, S. T. L. Tay, B. Y. P. Moy, and Y. Liu, Aerobic granulation: an innovative biotechnology for industrial wastewater treatment, In: *Proceeding of Sino-Singapore Joint Symposium on Environmental Protection Technologies and Environmental Hydraulics and First Asian Environmental Research Alliance Partnership Symposium*, January 06–09, 2003, pp. 215–222, Taipei, Taiwan (2003).
17. J. H. Tay, S. F. Yang, and Y. Liu, Hydraulic selection pressure-induced nitrifying granulation in sequencing batch reactors, *Applied Microbiology and Biotechnology*, 59, 332–337 (2002).
18. L. W. Hulshoff Pol, *The phenomenon of granulation of anaerobic sludge*, Ph.D. thesis, Wageningen Agricultural University, Wageningen, The Netherlands (1989).
19. N. Kosaric, R. Blaszczyk, L. Orphan, and J. Valladares, The characteristics of granules from upflow anaerobic sludge blanket reactors, *Water Research*, 24, 1473–1477 (1990).
20. C. M. M. Campos and G. K. Anderson, The effect of the liquid upflow velocity and the substrate concentration on the start-up and steady state period of lab-scale UASB reactors, *Water Science and Technology*, 25, 41–50 (1992).
21. J. H. Tay and Y. G. Yan, Influence of substrate concentration on microbial selection and granulation during start-up of upflow anaerobic sludge blanket reactors, *Water Environmental Research*, 68, 1140–1150 (1996).
22. J. T. C. Grotenhuis, J. B. van Lier, C. M. Plugge, A. J. M. Stams, and A. J. B. Zehnder, Effect of ethylene glycol-bis ((-aminoethyl ether)-N,N,-tetraacetic acid (EGTA) on stability and activity of methanogenic granular sludge, *Applied Microbiology and Biotechnology*, 36, 109–114 (1991).
23. L. Morvai, P. Mihaltz, and L. Cazko, The kinetics basis of a new start-up method to ensure the rapid granulation of anaerobic sludge, *Water Science and Technology*, 25, 113–122 (1992).
24. J. Quarmby and C. F. Forster, An examination of structure of UASB granules, *Water Research*, 29, 2449–2545 (1995).

25. P. A. Alphenaar, A. Visser, and G. Lettinga, The effect of liquid upflow velocity and hydraulic retention time on granulation in UASB reactors treating wastewater with a high-sulphate content, *Bioresource Technology*, 43, 249–258 (1993).
26. Y. Arcand, S. R. Guitot, M. Desrochers, and C. Chavarie, Impact of the reactor hydrodynamics and organic loading on the size and activity of anaerobic granules, *Journal of Chemical Technology and Biochemistry*, 56, 23–35 (1994).
27. V. O'Flaherty, P. N. Lens, D. de Beer, and E. Colleran, Effect of feed composition and upflow velocity on aggregate characteristics in anaerobic upflow reactors, *Applied Microbiology and Biotechnology*, 47, 102–107 (1997).
28. Y. Liu and J. H. Tay, Detachment forces and their influence on the structure and metabolic behavior of biofilms, *World Journal of Microbiology and Biotechnology*, 17, 111–117 (2001).
29. Y. Liu and J. H. Tay, The essential role of hydrodynamic shear force in the formation of biofilm and granular sludge, *Water Research*, 36, 1653–1665 (2002).
30. M. C. M. van Loosdrecht, D. Eikelboom, A. Gjaltema, A. Mulder, L. Tijhuis, and J. J. Heijnen, Biofilm structures, *Water Science and Technology*, 32, 35–43 (1995).
31. H. T. Chang, B. E. Rittmann, D. R. Amar, O. Ehrlinger, and Y. Lesty, Biofilm detachment mechanisms in a liquid fluidized bed, *Biotechnology and Bioengineering*, 38, 499–506 (1991).
32. W. K. Kwok, C. Picioreanu, S. L. Ong, M. C. M. van Loosdrecht, W. J. Ng, and J. J. Heijnen, Influence of biomass production and detachment forces on biofilm structures in a biofilm airlift suspension reactor, *Biotechnology and Bioengineering*, 58, 400–407 (1998).
33. H. S. Shin, K. H. Lim, and H. S. Park, Effect of shear stress on granulation in oxygen aerobic upflow sludge bed reactor, *Water Science and Technology*, 26, 601–605 (1992).
34. J. H. Tay, Q. S. Liu, and Y. Liu, The effects of shear force on the formation, structure and metabolism of aerobic granules, *Applied Microbiology and Biotechnology*, 57, 227–233 (2001).
35. M. J. Vieira, L. F. Melo, and M. M. Pinheiro, Biofilm formations: hydrodynamic effects on internal diffusion and structure, *Biofouling*, 7, 67 (1993).
36. S. Wasche, H. Horn, and D. C. Hempel, Mass transfer phenomena in biofilm systems, *Water Science and Technology*, 41, 357 (2000).
37. W. R. Ross, The phenomenon of sludge pelletization in the anaerobic treatment of a maize processing waste, *Water SA*, 10, 197–204 (1984).
38. B. E. Christensen, The role of extracellular polysaccharides in biofilms, *Journal of Biotechnology*, 10, 181–202 (1989).
39. J. H. Tay, Q. S. Liu, and Y. Liu, The role of cellular polysaccharides in the formation and stability of aerobic granules, *Letters in Applied Microbiology*, 33, 222–226 (2001).
40. F. Trinet, R. Heim, D. Amar, H. T. Chang, and B. E. Rittmann, Study of biofilm and fluidization of bioparticles in a three-phase fluidized-bed reactor, *Water Science and Technology*, 23, 1347 (1991).
41. A. Ohashi and H. Harada, Adhesion strength of biofilm developed in an attached-growth reactor, *Water Science and Technology*, 29, 10 (1994).
42. M. J. Chen, Z. Zhang, and T. R. Bott, Direct measurement of the adhesive strength of biofilms in pipes by micromanipulation, *Biotechnology Techniques*, 12, 875–880 (1998).
43. J. T. C. Grotenhuis, J. C. Kissel, C. M. Plugge, and A. J. B. Zehnder, Role of substrate concentration in particle size distribution of methanogenic granular sludge in UASB reactors, *Water Research*, 25, 21–25 (1991).
44. J. E. Schmidt and B. K. Ahring, Effects of magnesium on thermophilic acetate-degrading granules in upflow anaerobic sludge blanket (UASB) reactors, *Enzyme Microbial Biotechnology*, 15, 304–310 (1993).

45. C. F. Shen, N. Kosaric, and R. Blaszczyk, The effect of selected heavy metals (Ni, Co and Fe) on anaerobic granules and their extracellular polymeric substance (EPS), *Water Research*, 27, 25–33 (1993).
46. L. W. Hulshoff Pol, W. J. de Zeeuw, C. T. M. Velzeboer, and G. Lettinga, Granulation in UASB reactors, *Water Science and Technology*, 15, 291–304 (1983).
47. E. M. Mahoney, L. K. Varangu, W. L. Cairns, N. Kosaric, and R. G. E. Murray, The effect of calcium in microbial aggregation during UASB start-up, *Water Science and Technology*, 19, 249–260 (1987).
48. J. T. C. Grotenthuis, J. van Lier, and C. M. Plugge, Effect of calcium removal on size and strength of methanogenic granules, In: *Poster-Papers, Fifth International Symposium on Anaerobic Digestion*, 22–26 May 1988, A. Tilche and A. Rozzi (eds.), pp. 117–120, Bologna, Italy (1988).
49. H. L. Jiang, J. H. Tay, Y. Liu, and S. T. L. Tay, Ca^{2+} augmentation for enhancement of aerobically grown microbial granules in sludge blanket reactors, *Biotechnology Letters*, 25, 95–99 (2003).
50. K. C. Marshall and R. H. Gruickshank, Cell surface hydrophobicity and the orientation of certain bacteria at interfaces, *Archives of Microbiology*, 91, 29–40 (1973).
51. B. Del Re, B. Sgorbati, M. Miglioli, and D. Palenzona, Adhesion, autoaggregation and hydrophobicity of 13 strains of *Bifidobacterium longum, Letters in Applied Microbiology*, 31, 438–442 (2000).
52. J. H. Tay, Q. S. Liu, and Y. Liu, Characteristics of aerobic granules grown on glucose and acetate in sequential aerobic sludge blanket reactors, *Environmental Technology*, 23, 931–936 (2002).
53. J. R. Zhu and P. A. Wilderer, Effect of extended idle conditions on structure and activity of granular activated sludge, *Water Research*, 37, 2013–2018 (2003).
54. J. H. Tay, V. Ivanov, S. Pan, and S. T. L. Tay, Specific layers in aerobically grown microbial granules, *Letters in Applied Microbiology*, 34, 254–257 (2002).
55. S. T. L. Tay, V. Ivanov, S. Yi, W. Q. Zhuang, and J. H. Tay, Presence of Anaerobic Bacteroides in Aerobically Grown Microbial Granules, *Microbial Ecology*, 44, 278–285 (2002).
56. S. K. Toh, S. T. L. Tay, B. Y. P. Moy, V. Ivanov, and J. H. Tay, Size-effect on the physical characteristics of the aerobic granules in a SBR, *Applied Microbiology and Biotechnology*, 60, 687–695 (2003).
57. J. H. Tay, S. T. L. Tay, V. Ivanov, S. Pan, and Q. S. Liu, Biomass and porosity profile in microbial granules sued for aerobic wastewater treatment, *Letters in Applied Microbiology*, 36, 297–301 (2003).
58. S. Yi, J. H. Tay, A. M. Maszenan, and S. T. L. Tay, A culture-independent approach for studying microbial diversity in aerobic granules, In: *Proceedings of IWA Asian Environmental Technology Conference*, October 30–November 1, 2001, pp. 486–494 (2001).
59. J. H. Pringle and M. Fletcher, Influence of substratum wettability on attachment of fresh bacteria to solid surface, *Applied and Environmental Microbiology*, 45, 811–817 (1983).
60. J. H. Tay, H. L. Xu, and K. C. Teo, Molecular mechanism of granulation. I: H^+ translocation-dehydration theory, *Journal of Environmental Engineering*, 126, 403–410 (2000).
61. Y. Liu, S. F. Yang, Q. S. Liu, and J. H. Tay, The Role of cell hydrophobicity in the formation of aerobic granules, *Current Microbiology*, 46, 270–274 (2003).
62. D. H. Applegate and J. D. Fryers, Effects of carbon and oxygen limitations and calcium concentration on biofilm removal processes, *Biotechnology and Bioengineering*, 37, 17–25 (1991).
63. J. E. Schmidt and B. K. Ahring, Extracellular polymers in granular sludge from different upflow anaerobic sludge blanket (UASB) reactors, *Applied Microbiology and Biotechnology*, 42, 457–462 (1994).

64. H. Harada, G. Endo, Y. Tohya, and K. Momomoi, High rate performance and its related characteristics of granulated sludges in UASB reactors treating various wastewaters, In: *Poster papers of the 5th International symposium on Anaerobic digestion*, 22–26 May 1988, A. Tilche, A. Rozzi (eds.), pp. 1011–1022, Bologna, Italy (1988).
65. P. Vandevivere and D. L. Kirchman, Attachment stimulates exopolysaccharide synthesis by a bacteria, *Applied and Environmental Microbiology*, 59, 3280–3286 (1993).
66. P. N. Danese, L. A. Pratt, and R. Kolter, Exopolysaccharide production is required for development of *Escherichia coli* K-12 biofilm architecture, *Journal of Bacteriology*, 182, 3593–3596 (2000).
67. A. Lodi, C. Solisoio, A. Converti, and M. Del Borghi, Cadmium, Zinc, Copper, Silver and Chromium(III) removal from wastewaters by *Sphaerotilus natans, Bioprocess Engineering*, 19, 197–203 (1998).
68. J. Taniguchi, H. Hemmi, K. Tanahashi, N. Amano, T. Nakayama, and T. Nishim, Zinc biosorption by a zinc-resistant bacterium, *Brevibacterium sp*. Strain HZM-1, *Applied Microbiology and Biotechnology*, 54, 581–588 (2000).
69. E. Valdman and S. G. F. Leite, Biosorption of Cd, Zn and Cu by *Saragssum sp*. waste biomass, *Bioprocess Engineering*, 22, 171–173 (2000).
70. Y. Lim, M. C. Lam, and H. H. P. Fang, Adsorption of heavy metals by EPS of activated sludge, *Water Science and Technology*, 43, 59–66 (2002).
71. Y. Liu, S. F. Yang, S. F. Tan, Y. M. Lin, and J. H. Tay, Aerobic granules: a novel zinc biosorbent, *Letters in Applied Microbiology*, 35, 548–551 (2002).
72. Y. Liu, S. F. Yang, H. Xu, K. H. Woon, Y. M. Lin, and J. H. Tay, Biosoption kinetics of cadmium (II) on aerobic granular sludge, *Process Biochemistry*, 38, 995–999 (2003).
73. J. R. Zhu and C. X. Liu, Cultivation and physico-chemical characteristics of granular activated sludge in alternation of anaerobic/aerobic process, *Chinese Journal of Environmental Science*, 20, 38–41 (1999).
74. J. H. Tay, S. Pan, Y. X. He, and S. T. L. Tay, Effect of organic loading rate on aerobic granulation. Part I: reactor performance, *Journal of Environmental Engineering*, 130, 1094–1101 (2004).
75. J. H. Tay, S. Pan, Y. X. He, and S. T. L. Tay, Effect of organic loading rate on aerobic granulation. Part II: Characteristics of aerobic granules, *Journal of Environmental Engineering*, 130, 1415–1423 (2004).
76. Y. Liu, S. F. Yang, J. H. Tay, Q. S. Liu, L. Qin, and Y. Li, Cell hydrophobicity is a triggering force of biogranulation, *Enzyme Microbial Technology*, 34, 371–379 (2004).

4
Membrane Bioreactors

Lawrence K. Wang and Ravinder Menon

CONTENTS

Abstract Membrane bioreactor (MBR) is a biochemical engineering process involving the use of both (a) a suspended growth bioreactor for biochemical reactions (such as fermentation, bio-oxidation, nitrification, and denitrification); and (b) a membrane separator for sequent solid–liquid separation. In a chemical engineering fermentation process, the solids may be yeasts and the liquid may be an alcohol. In an environmental engineering process, the solids may be activated sludge, and the liquid may be the biologically treated wastewater (WW).

Practically speaking, the membrane separator replaces clarifier, such as sedimentation or dissolved air flotation in a conventional activated sludge (CAS) process system. The membrane module can be either submerged in the activated sludge bioreactor, or situated outside the activated sludge bioreactor. This chapter introduces historical development, engineering applications, various MBR process systems, design considerations and practical environmental engineering applications, such as treatment of dairy industry wastes, landfill leachate, coffee industry wastes, and cosmetics industry wastes.

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_4 © Humana Press, New York, NY 2009

Key Words Activated sludge • bio-oxidation • coffee industry waste • cosmetics industry waste • dairy industry waste • denitrification • design • fermentation • landfill leachate • MBR systems • membrane bioreactor • membrane process • nitrification • suspended growth.

1. INTRODUCTION

1.1. General Introduction

With increasing pressures worldwide on existing water resources due to increases in human population and activity, reuse, and conservation of water resources assumes a very high priority. The membrane bioreactor (MBR) is a tool that has the potential to help industries (and municipalities) manage their water resources better. The MBR is an innovative wastewater treatment (WWT) technology, based on proven processes of activated sludge biological treatment and membrane separation (1–50). The system has been implemented in several full-scale industrial and municipal applications. The synergistic combination of enhanced biological treatment and membrane filtration produces a treated effluent quality, which is not merely excellent, but very reliable. This provides the opportunity to facilities to recycle/reuse part or all the treated effluent, thereby reducing costs for fresh water and water treatment on the one hand, and reducing sewer surcharge (for pretreatment facilities), on the other. This chapter discusses some basic aspects of design of MBRs, and presents some full-scale examples of its application.

1.2. Historical Development

1.2.1. Membrane Processes

The chemical engineering processes involving the use of membranes for phase separation are termed "membrane processes." The phases include solid phase (suspended solids [SS], dissolved solids, etc.), liquid phase (water, ethanol, chloroform, etc.), gas phase (air, nitrogen, oxygen, etc.). A membrane is a porous filtration medium, which can be cationic, anionic, or nonionic in nature, and acts as a barrier to prevent mass movement of selected phases, but allows passage of remaining phases. The main applications of the membrane processes are processing water and wastewater streams (1). Recently membrane processes have been used for purification of gas streams (2).

The membrane processes include at least five main subcategories for processing water and wastewater (1).

1.2.1.1. MICROFILTRATION (MF)

Microfiltration is a pressure-filtration process for the separation of suspended solids in the particle size-range of about 0.08 to 10 μm. The primary function affecting solids separation from water is the size of suspended solids (SS). The hydraulic pressure (transmembrane pressure) applied in microfiltration (MF) is about 1 to 2 bars, or 15 to 20 psig, primarily for overcoming resistance of the "cake." (1 micron = 1 μm = 0.00004 in. = 10,000 Å) (1).

1.2.1.2. Ultrafiltration (UF)

Ultrafiltration is another pressure filtration process for the separation of macromolecular solids in the particle size range of about 0.01 to 0.1 μm. The primary factor affecting solids separation from water relies on the size of macromolecular solids. The hydraulic pressure required by ultrafiltration (UF) for overcoming hydraulic resistance of the polarized macromolecular layer on the membrane surface is about 1 to 7 bars (1).

1.2.1.3. Nanofiltration (NF)

Nanofiltration (NF) membranes are multiple-layer thin-film composites of polymer consisting of negatively charged chemical groups, and are used for retaining molecular solids (such as sugar), and certain multivalent salts (such as magnesium sulfate), but passing substantial amounts of most monovalent salts (such as sodium chloride), at an operating pressure of about 14 bars or 200 psig. Both molecular diffusivity and ionic charge play important roles in the separation process. The sizes of molecular solids and multivalent salts to be rejected by NF are normally in the range of 0.0005 to 0.007 μm (1).

1.2.1.4. Reverse Osmosis (RO)

Reverse osmosis (RO) membranes are mainly made of cellulose acetate with the pore sizes of about 5 to 20 Å, and are used for rejecting salts (as high as 98%) and organics (as high as 100%), at an operating pressure of about 20 to 50 bars or 300 to 750 psig. The hydraulic pressure (through a pump) is used to provide the driving force for permeation, or for overcoming the chemical potential difference between the concentrate and the permeate, expressed in terms of the osmotic pressure. The sizes of molecular solids and salts (multivalent as well as monovalent) to be rejected by RO are normally in the range of 0.00025 to 0.003 μm (1).

1.2.1.5. Electrodialysis (ED)

Electrodialysis (ED) uses voltage or current as the driving force to separate ionic solutes. The size of ionic solutes to be rejected or separated by ED are normally in the range of 0.00025 to 0.08 μm, depending on the pore size of ED membranes. EDR is the electrodialysis reversal (or reverse electrodialysis) process, which is similar to ED, but its cathodes and anodes can be reversed for automatic cleaning during operation (1).

Figure 4.1 illustrates the relationships among microfiltration (MF), ultrafiltration (UF), nanofiltration (NF) and reverse osmosis (RO) and conventional sand filtration process. Figure 4.2 shows the effects of MF, UF, NF, and RO on separation of suspended particles, macromolecules, sugar and salts.

In this chapter, mainly MF and UF are related to membrane bioreactors. RO is an effective post-treatment unit to MBR (*see* Section 5.2).

1.2.2. Physical–Chemical Pretreatment before Membrane Process

Theoretically and technically speaking, the membrane process alone will be feasible for purifying water or wastewater. For membrane process operation, build-up of a layer on the surface of the membrane and deposition of foulants within the membrane pore structure attributable to high solution concentrations are the major mechanisms responsible for

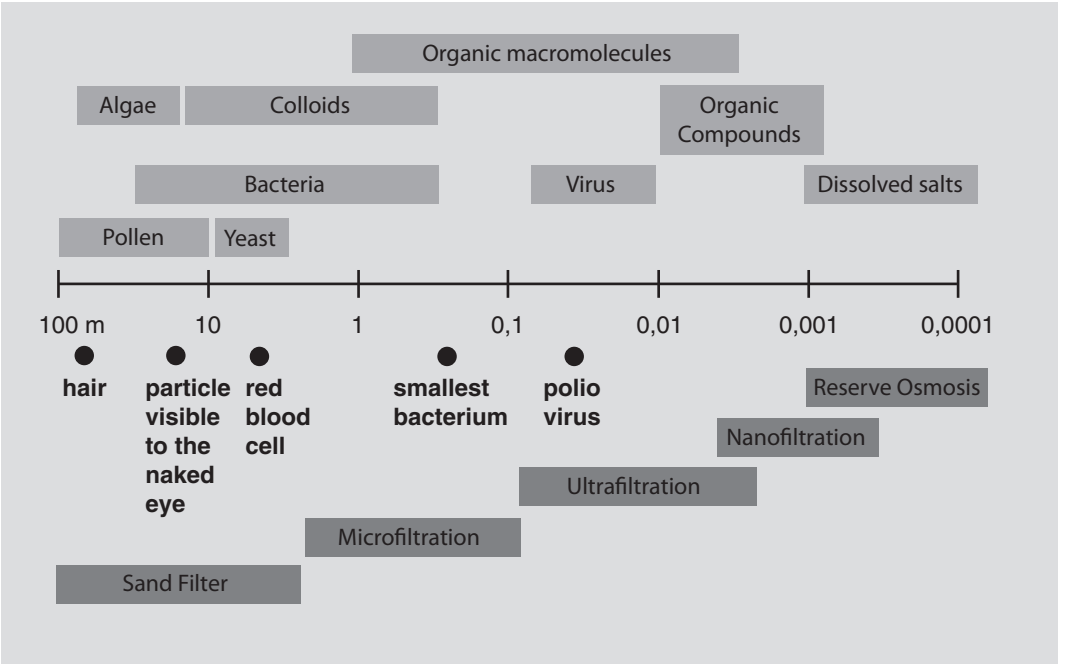

Fig. 4.1. Particle size and separation processes.

membrane flux decline (1). Membrane fouling requires frequent chemical cleanings, and in the worst case, membrane replacement.

Accordingly, the membrane process alone is mainly used for water purification when the quality of an influent raw water is good.

For potable water or industrial water treatment using membrane processes, physical–chemical pretreatment before membrane reactor will significantly prolong the membrane life, in turn, reduce the treatment cost (3). Typical physical–chemical pretreatment processes include: pH adjustment, chemical coagulation, clarification, sand filtration or cartridge cloth filtration (3). An ultra-low pressure drop oleophilic filter is also effective (41).

Again, a membrane process reactor with physical–chemical pretreatment is technically feasible but not economically feasible, for treatment of industrial waste or domestic sewage, due to high organic load in the influent streams.

1.2.3. Biological Pretreatment Prior to Membrane Process

Traditionally, biological treatment processes, such as activated sludge, tickling filters, lagoons, fluidized-bed reactors, rotating biological contactors, sequencing batch reactors (SBRs), adopt either sedimentation clarification, or dissolved air flotation clarification for solids–liquid separation, and for microorganisms (sludge) return (4–11).

When a biological process is used in conjunction with a membrane reactor (either microfiltration or ultrafiltration), the entire process system becomes a membrane bioreactor. The

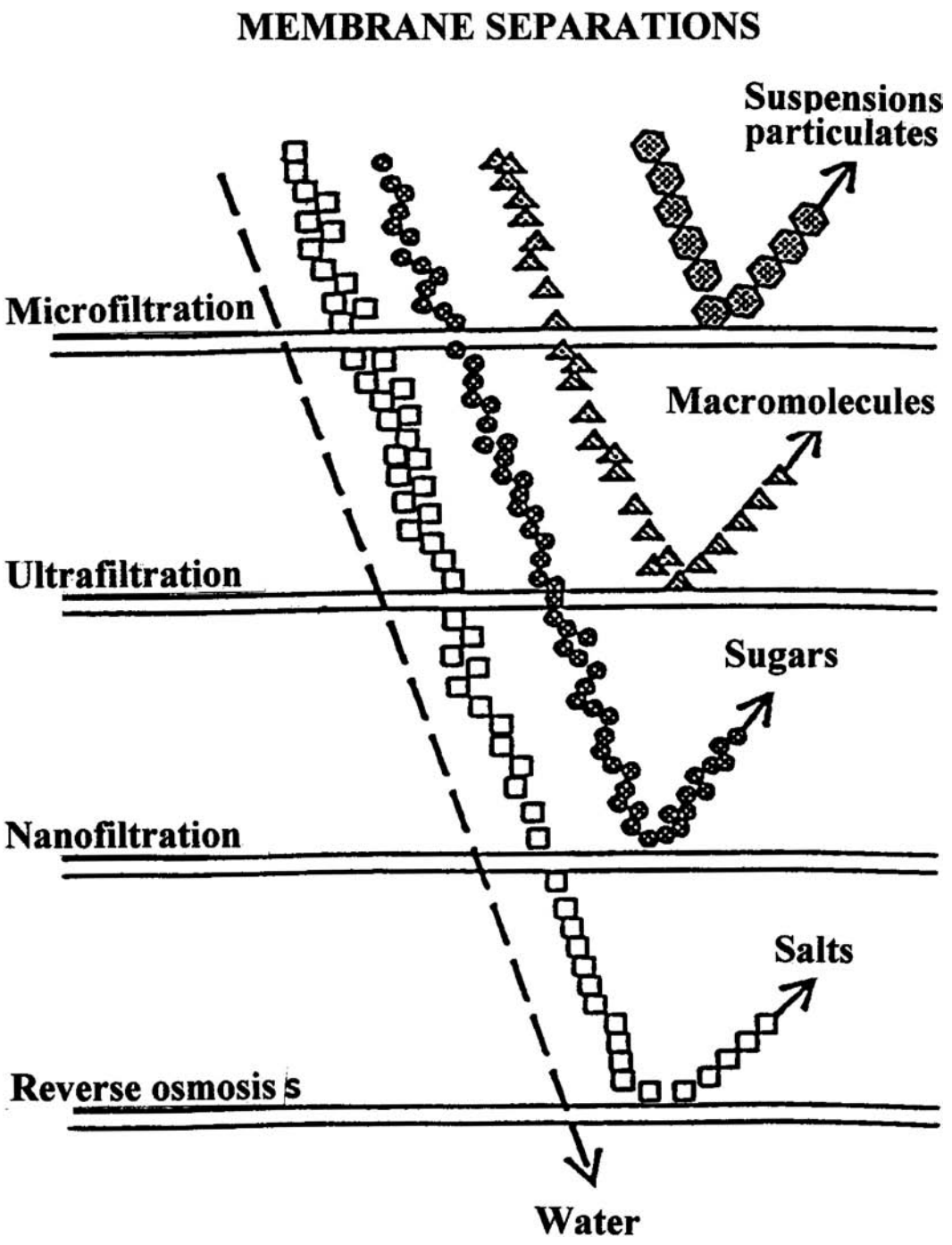

Fig. 4.2. Separation capabilities of MF, UF, NF, and RO.

membrane bioreactor (either microfiltration or ultrafiltration) will accomplish the following three tasks:

1. Solids–water separation (clarification)
2. Sludge return (microorganisms return)
3. Tertiary wastewater treatment, capable of disinfection, nutrients removal, metals removal, and toxic organics removal.

When treating wastewater streams at same flow, biological treatment processes are usually more cost-effective than comparable physical–chemical processes.

Biological processes can: (a) adopt suspended growth reactors, attached growth reactors, or both, (b) be operated as a continuous process system or a sequencing batch reactor process system, and (c) be controlled under aerobic, anoxic, or anaerobic conditions for biochemical treatment of wastewater. The theory and principles of biological treatment processes can be found elsewhere (4–10, 49, 51, 52), and are beyond the scope of this chapter. In general, biological processes can accomplish carbonaceous oxidation (aerobic), nitrification (aerobic), denitrification (anoxic/anaerobic), digestion (aerobic or anaerobic), phosphorus removal (aerobic/anoxic), and methane production (anaerobic).

There are many kinds of MBR systems, which will be introduced in Section 1.3, MBR Research and Engineering Applications.

Sequencing batch reactor (SBR) can also adopts membrane modules forming a SBR–MBR batch process system for wastewater treatment (51).

Only the common MBRs consisting of activated sludge aeration basins and microfiltration or ultrafiltration will be presented and discussed in this chapter in detail.

Although MBRs are mainly used for wastewater treatment, they can also be used for potable water treatment, aiming at nitrogen removal. (50)

1.2.4. Physical–Chemical–Biological Pretreatment Before Membrane Process

It has been known that many nonbiodegradable or toxic pollutants can not be removed by biological processes. On the other hand, many pollutants can not be chemically coagulated or adsorbed (10, 12).

For treatment of certain heavily polluted wastewater streams, it may be necessary to adopt combined physical–chemical–biological pretreatment before a membrane process reactor.

1.3. Membrane Bioreactors Research and Engineering Applications

In this section, various membrane bioreactors will be reviewed and discussed, although the remaining sections of this chapter will introduce and discuss only the most common MBRs, which are well-established and practically applied to the environmental engineering field for pollution control.

MBRs consisting of an aerobic or anaerobic reactor with suspended biomass and membranes for liquid–solid separation are now the mainstream environmental engineering processes for water and wastewater treatment. MBRs have been used for treatment of municipal and industrial wastewater (13), and for reclamation of municipal wastewater for potential reuse in public water supplies (14).

With respect to treatment efficiency and system stability, MBRs have several advantages over conventional processes:

1. With complete solids–liquid separation by the membrane, high biomass concentrations and relatively short reaction times are possible (13–21).
2. MBRs can produce a clear final effluent regardless of hydraulic retention time and without concerns of biomass settleability characteristics (15).
3. Biological nitrogen removal is also possible in an intermittently aerated single-stage MBR (16).

A MBR with powdered activated carbon (PAC) addition was applied for drinking water treatment to remove nitrate, natural organic matter, and pesticides and to disinfect the water (22). Also, the addition of PAC to the activated-sludge process with attached microbial growth on the PAC enhanced membrane permeability. The flux enhancement could be attributed to the development of dense floc particles around the PAC (23, 52).

Brindle and Stephenson (13) studied three generic membrane processes within bioreactors for wastewater treatment, solids separation and recycle, bubble-less aeration, and priority organic pollutants extraction. Commercial aerobic and anaerobic MBRs are already in use producing a high-quality effluent at high organic loading rates. However, bubble-less aeration and extractive MBRs are still in development.

Dollorer and Wilderer (24) compared oxygenation by bubbling and via a silicone rubber, bubble-free membrane system in sequencing batch biofilm reactors (SBBRs). The clay-packed SBBRs achieved 68% dissolved organic compounds removal from hazardous waste landfill leachate with a 12 h cycle. The bubble-free SBBR emitted less biodegradable volatile organics than the bubbled system.

Livingston et al. (25) used an extractive membrane bioreactor (EMB) to remove a range of toxic organic compounds from the chemical industry, achieving more than 99% removal with a wastewater reactor contact time of less than 30 minutes. The removal efficiency was modeled, and a new EMB configuration was discussed. Data on the effect of biofilms on membrane mass transfer were shown. In additional work, this group demonstrated that addition of sodium chloride to the biomedium increases the maintenance energy requirement of the degradative organisms and resulted, in a carbon-limited situation, in reduction of biofilm growth. Organic substrate flux remained high under reduced biofilm growth conditions (26).

Cote et al. (27) discovered that when a submerged membrane was placed in an aeration tank for municipal wastewater treatment with an anoxic-aerobic process, total Kjeldahl nitrogen (TKN) removal efficiencies were greater than 69% and 94% at mixed liquor suspended solids (MLSS) concentrations of 15,000 and 25,000 mg/L, respectively. Further studies on aeration strategies to optimize nitrogen removal designs are needed. The application of membranes to biological wastewater treatment is limited by membrane fouling and high energy consumption. Back-flushing with permeate or air in a cross-flow membrane coupled to a biological reactor has been used to reduce membrane fouling (22, 28, 29). An improvement in flux rates compared to that for operations without back-flushing was reported.

Air contaminated with trichloroethylene (TCE) was passed through microporous hollow fibers in a hollow-fiber membrane bioreactor whereas an oxygen-free nutrient solution was

recirculated through the shell side of the membrane module. A removal efficiency of 30% was achieved at inlet TCE concentrations of 20 ppmv and a 36-s gas phase residence time (30).

A bioreactor was developed by Clapp et al. (31) using silicone tubing with an attached methanotrophic biofilm to treat TCE-contaminated waste streams. The reactor was developed to overcome the low solubility of methane, competitions between methane and TCE, the lack of NADH regeneration in the presence of TCE only, and the cytotoxic products of TCE metabolism.

Many other techniques such as formation of a dynamic membrane, precoat, or hydrophobic skin layers atop the membrane, have been introduced to reduce fouling in cross-flow MBRs, but these are still in an early stage of evaluation. Some researchers using cross-flow MBRs have reported that the pumping shear stress caused biological floc break-up, leading to severe flux decline in long-term operations caused by the small flocs forming a denser biomass cake layer on the membrane. Additionally, continuous recycling of mixed liquor in a cross-flow MBR requires a relatively large amount of energy (32–35).

Yamamoto et al. (36) studied an alternative to a cross-flow membrane operation using a submerged membrane with permeate removal by vacuum suction. The power consumption per unit volume of treated water was greatly reduced by eliminating the circulation pump, but the permeate flux was reduced to an impractical low level of less than 2 L/m^2 h. The energy consumption associated with filtration in these new submerged membrane reactors was at a substantially low level of 0.2 to 0.4 kWh/m^3 treated compared to the relatively high energy consumption (2 to 10 kWh/m^3) with circulation loops (27).

Performance of a sequencing batch reactor using a membrane for effluent filtration was investigated by Choo and Stensel (37). In terms of chemical oxygen demand (COD) removal, nitrogen removal, and membrane permeability during long-term continuous operation treating synthetic wastewater, the reactor was operated with six 4-hour cycles per day consisting of 0.2, 2.0, and 1.5 hours for fill, aeration, and effluent filtration-idle respectively. Minimal solids wasting occurred for the first 10 months of operation followed by an 8-day solids retention time (SRT) for the final 1.5 months. Membrane fouling was controlled by back-washing with aeration for 10 min during each cycle. A stable permeate flux of approximately 0.34 (L/m^2 h)/kPa, or (34 (L/m^2 h)/bar) was achieved and was independent of mixed liquor suspended solids concentrations from 700 to 10,000 mg/L. The reactor effluent turbidity averaged less than 0.20 NTU, and more than 98% COD removal occurred. Nitrogen removal efficiency ranged from 87% to 93% through biological nitrification and denitrification. Most of the nitrate was removed during the mixed and unaerated fill period, but a significant amount of nitrogen was removed by simultaneous nitrification-denitrification (SNDN) during aeration at dissolved oxygen (DO) concentrations less than 3.0 mg/L.

A new membrane separation process developed by Osmotek of Corvallis, Oregon, USA, is moving out of the pilot testing phase, and is available for a variety of applications, such as treating wastewater and landfill leachate, according to the US Department of Energy (DOE), which helped develop the technology (38).

The technology called direct osmosis concentration (DOC), is a cold temperature membrane process that separates waste streams in a low-pressure environment. DOC uses salt

brine as an osmotic agent to treat wastewater on board US Navy vessels. The technology also has been shown to remove 95% of the water from leachate with little maintenance (38).

Anaerobic wastewater treatment processes have become increasingly popular for treating industrial effluents, especially those containing high levels of fermentative products. Because of their ability to withhold slow-growing bacteria, anaerobic membrane bioreactors have generated increased interest in recent years.

Beaubien et al. (20) used a 1.5 m^3 anaerobic MBR pilot plant to treat condensate from a distillery to evaluate the possibility of recycling and reusing the treated effluent in alcoholic fermentations. Consisting mostly of acetate, propionate, and ethanol with a mean chemical oxygen demand (COD) of 3,000 mg/L, distillery condensates are particularly suitable for anaerobic treatment. Following a 5-day biological anaerobic MBR start-up period, during which removal efficiency increased from 40% to 80%, satisfactory performance of the anaerobic membrane bioreactor was obtained. More than 75% of the applied load was removed. The suitability of the treated effluent for reuse in alcoholic fermentations was evaluated by comparing the alcohol concentration obtained using treated and untreated effluents, and water as process-dilution fluid in fermentations. The results clearly show that the untreated effluent significantly inhibits the fermentative organisms, whereas treated effluent does not induce a noticeable inhibition of alcoholic fermentation.

2. MBR PROCESS DESCRIPTION

2.1. Membrane Bioreactor with Membrane Module Submerged in the Bioreactor

This type of MBR process uses the same biological wastewater treatment as conventional activated sludge (CAS), but provides tertiary treatment with far fewer unit processes. Aeration, secondary clarification, and filtration (without the need for coagulation/flocculation) occur within a single bioreactor (shown in Fig. 4.3), rather than in separate basins.

The MBR process uses hollow-fiber microfiltration or ultrafiltration membranes. The membranes are bundled into "modules" and grouped together in "cassettes." The cassettes are connected by a header to a permeate (effluent) pump and are submerged in the bioreactor. In more recent configurations, the cassettes are submerged in separate tanks, for the ease of cleaning. The permeate pumps create a vacuum that pulls the effluent into the hollow-fiber membranes, but leaves the solids behind in the bioreactor. This eliminates the need for secondary clarification and return sludge pumping (18, 44).

Because activated sludge stays in the bioreactor, the concentration of MLSS is much higher (10,000 to 12,000 mg/L) than it would be in a conventional activated sludge process. The high MLSS concentrations facilitate treatment within a smaller bioreactor volume.

The hydraulic capacity of the MBR process is limited by the flow of water per unit area of the membrane surface. The average flow rate per unit area, or flux, for membranes that are used for WWT is typically 10 to 15 $gal/ft^2/day$.

This type of MBR process can be built from the ground up, or retrofitted into an existing CAS aeration basin, as shown in Fig. 4.3. In operation, air is supplied through coarse bubble diffusers at the base of the membrane cassettes to agitate and scour the membranes for

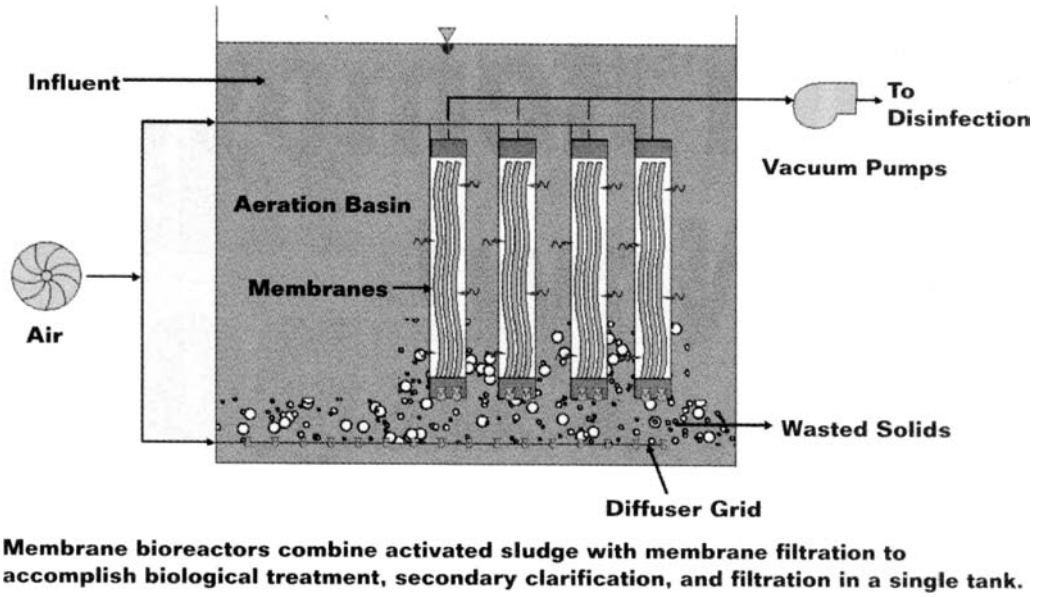

Fig. 4.3. MBR process system with membrane module submerged in the bioreactor.

cleaning and to provide oxygen for biological treatment. At regular intervals, automatic backwash (backpulse) cycles clean and restore permeability to the membranes. The coarse bubble diffusers used for membrane cleaning do not transfer oxygen efficiently, so fine bubble diffusers (or other means of aeration) are added to supply more air for treatment.

2.2. Membrane Bioreactor with Membrane Module Situated Outside the Bioreactor

This type of MBR process system is schematically shown in Fig. 4.4. Screened influent enters the bioreactor, where it is oxidized to remove organic pollution, as well as ammonia, if any. The mixed liquor from the bioreactor at an MLSS concentration ranging from 10 to 20 g/L is withdrawn and pumped through a crossflow membrane filtration module. The permeate from the membranes constitutes the treated effluent. The retentate stream representing concentrated biosolids is returned to the bioreactor. Excess biosolids are wasted from the bioreactor or from the return line.

It may be noted that due to the membranes acting as absolute barrier for solids, it is possible to accurately maintain the desired sludge age or solids retention time (SRT). Also the micro- or ultra-filtration membranes used for separation are capable of separating suspended and colloidal solids, organic macromolecules as well as micro-organisms from treated effluent. Figure. 4.1 and 4.2 illustrate this point.

The MBR system introduced in this section is based on an external or in-series configuration, where the membrane units follow and are situated outside the bioreactor. This helps keep the two processes separate, avoiding interferences and enabling individual optimization.

The complete separation of hydraulic and solids retention times provides optimum control of biological reactions, and greater reliability and flexibility in use (15, 20, 21). The MBR system typically uses high SRTs in the range of 60 to 100 days. The high SRT used helps in the development of slow-growing micro-organisms, such as nitrifying bacteria, as well as provides complete biodegradation of difficult-to-treat components such as organic macromolecules, which are retained by the membrane units, and kept in the system until biodegradation.

2.3. MBR System Features

The two most common types of MBR system are introduced in Sections 2.1 and 2.2 (*see* Figs. 4.3 and 4.4). Since both are very similar to each other, only one type (Section 2.2; Fig. 4.4) is described in detail in the remaining sections of this chapter. Entire MBR system shown in Fig. 4.4 includes, but is not limited to the screen, conditioning tank, bioreactor, pumps, and pipes. Special system features of the innovative MBR system (Fig. 4.4) include:

1. Very high quality bacteria-free effluent.
2. High organic loading loading (2 to 4 kg COD/m^3 day or 0.12 to 0.25 lb/ft^3 day).
3. High MLVSS (10,000 to 20,000 mg/L).
4. Efficient oxygen transfer.
5. Very high sludge ages used (30 to 100 days).
6. Thirty-five percent to 45% less excess biosolids (sludge) production.
7. Promoting growth of slow-growing bacteria.
8. Immune to filamentous and other bulking.

The working principle of the crossflow membrane filtration is illustrated in Fig. 4.5, based on one membrane module.

In operation, the mixed liquor from the bioreactor (*see* Fig. 4.4) passes a security filter, then enters the mixed liquor inlet of a membrane module, which consists of many bundles of membrane filters. Through the mixed liquor inlet, the mixed liquor enters the tube-type

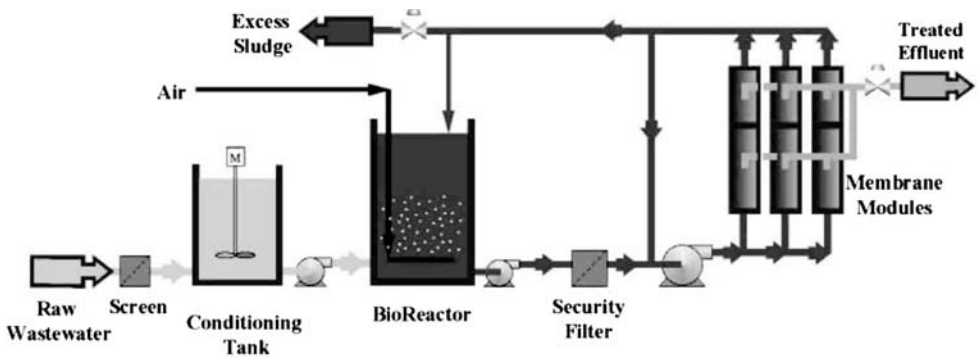

Fig. 4.4. MBR process system with membrane module situated outside the bioreactor.

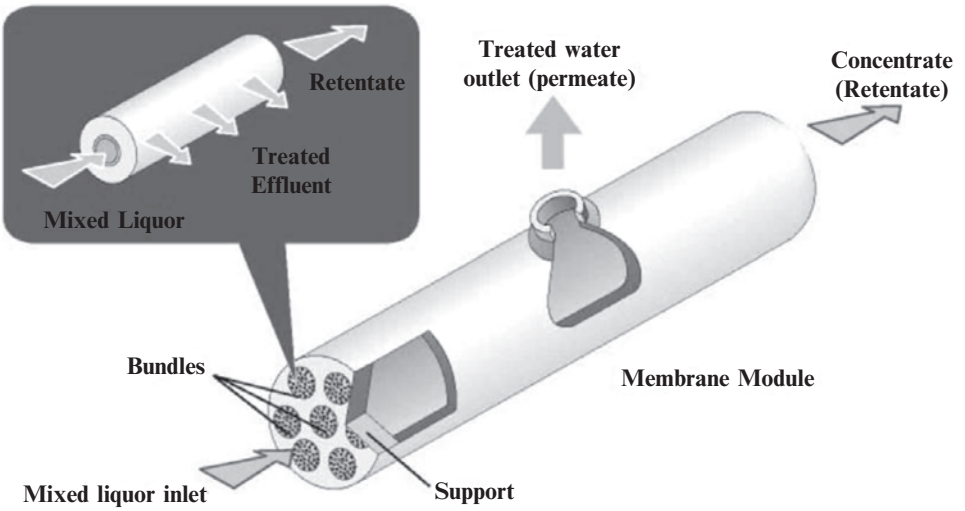

Fig. 4.5. Working principle of crossflow membrane filtration.

membrane filters where water-solid separation occurs under high pressure. The liquid portion of the mixed liquor is forced by pressure to pass through the tube-type membrane filter becoming the treated effluent, whereas the suspended solids remain becoming highly concentrated retentate (or concentrate). The suspended solids are mainly micro-organisms from the bioreactor having particle sizes larger than that of the membrane filter's pores.

The treated effluent (or treated water) as shown in Figs. 4.4 and 4.5, is discharged to a receiving water or reused.

The retentate from the membrane modules is partially returned to the bioreactor as return activated sludge (RAS), and partially wasted as excess sludge (Fig. 4.4).

Since the direction of mixed liquor flow inside of a tube-type membrane filter is perpendicular to the direction of treated effluent passing through the membrane, this flow pattern is called crossflow filtration—this is the first special membrane feature of the membrane filtration operation.

The second special membrane feature, created by ONDEO Degremont involves the use of ceramic membranes, which have very high corrosion resistance, and can be cleaned more efficiently, and are less prone to bio-fouling.

Availability of various pore sizes in ultrafiltration and microfiltration range (Figs. 4.1 and 4.2) is the third special membrane feature.

The fourth special membrane feature is that the membranes can be cleaned by CIP (clean-in-place) techniques, using crossflow filtration, and reverse backwash operation.

As discussed previously, a complete MBR system (Fig. 4.4) developed by ONDEO Degremont, is based on external or in-series configuration, where the membrane units follow, and are situated outside the bioreactor. This added special feature may keep the biological process in

the bioreactor, and the liquid–solid separation process (clarification) in the membrane module separate, avoiding interferences, and enabling individual process optimization.

2.4. Membrane Module Design Considerations

Membrane processes are characterized by two basic process parameters: (a) flux, which is the rate of transport of solvent or solution through the membrane; and (b) rejection, which is the degree of separation of a particular feed component.

There are five major variables that affect the two basic process parameters: (a) driving force in terms of applied transmembrane pressure and/or electric voltage/current; (b) flow velocity which affects turbulence and mass transfer coefficient; (c) process water temperature which has effects on physical properties such as density, viscosity, diffusivity, osmotic pressure, surface tension and others; (d) feed stream characteristics in terms of particle concentration, particle size, viscosity, molecular weight, molecular configuration, ionic charges, and fouling potential; and (e) membrane module in terms of materials, pore sizes, membrane configuration, membrane ionic charges, and feed compatibility (1).

There are basically six different designs of membrane modules: (a) tubular modules with channel diameters greater than 3 mm; (b) hollow fiber or capillary modules made of self-supporting tubes, usually 2 mm or less in internal diameters; (c) plate modules; (d) spiral-wound modules; (e) pleated sheet modules; and (f) rotary modules. The latter four module designs use flat sheets of membrane in various configurations (1).

In selecting a particular membrane module and a particular membrane process, the major criteria are: (a) feed stream characteristics, which affect the biocompatibility of the membranes; (b) flux requirements, which are controlled by the volumetric rate of a feed stream; (c) rejection requirements, which decide the process objectives and treatment efficiencies; and (d) cost requirements, which are affected by energy consumption, membrane replacement cost and operating and cleaning costs.

Biocompatibility of the membrane relates to the interaction between the membrane module and the feed stream. Major biocompatibility factors include: (a) stability to extremes in temperature, pressure, and pH, especially under cleaning and sanitizing conditions; (b) membrane–solute interactions, which affect the rate of fouling, cleaning, yields, and rejection of individual feed substances; and (c) acceptability of the membrane as a contact material for the final product, which essentially implies using membrane materials that are inert and do not leach out any toxic substances from the membrane into the final product. In this regard, there are new generations of membranes, made of expensive inorganic materials, such as ceramics, stainless steel, carbon–zirconia, etc.

MF membranes are made of a wide range of inorganic materials (such as alumina, zirconia–carbon composites, carbon–carbon composites, ceramics, stainless steel, silica, etc.) and natural and synthetic polymers (such as polypropylene, polycarbonates, polysulfone, polyvinylchloride, PVC copolymer, cellulose esters, cellulose acetate, etc.) (1)

UF membranes are mainly made of polysulfone-type materials (such as polyether sulfone, polyphenyl sulfone, sulfonated polysulfone, etc.) although they are also available in a wide range of organic materials (such as PVC copolymer, cellulose acetate, etc.) and inorganic materials (such as ceramic composites, stainless steel, etc.).

Most NF membranes are multiple-layer thin-film composites of synthetic polymers. The active NF membrane layer usually consists of negatively charged chemical groups. NF membranes are of porous filter media with an average pore diameter of 2 nm. The nominal molecular weight cutoff ranges from 100 to 200. The active NF membrane layer can be made of polyamide, polyvinyl alcohol, sulfonated polysulfone, and sulfonated polyether sulfone. Salt rejection by NF membranes is mainly due to electrostatic interaction between the ions and the NF membrane. Rejection of neutral substances is by size.

Cellulose acetate and derivatives are widely used as the RO membrane, despite their real and perceived limitations. Thin-film composite membranes containing a polyamide separating barrier on a polysulfone or polyethylene supporting layer, generally give better performance for RO applications with regard to temperature and pH stability and cleanability, but have almost zero chlorine resistance. In general, these thin-film composite membranes will be the material of choice for RO applications, unless there is a specific fouling problem with these membranes.

There are four types of membrane equipment: tubular membrane modules, hollow-fiber membrane modules, plate membrane modules, and spiral-wound membrane modules. Each design has its own special applications, advantages, and disadvantages (1).

The large-bore tubular membrane modules are suitable for food streams with high concentrations of suspended solids such as citrus juices and animal waste streams, even though the tubular membrane modules have the lowest packing densities and highest energy consumption among all the modules. The tubular designs with ceramic inorganic membranes are frequently used in the food processing industries.

The hollow-membrane modules have extremely high packing density (surface area to volume ratios) and comparatively low energy consumption, and are suitable for comparatively clean feed streams with low concentrations of suspended solids and macromolecules. Certain macromolecules display non-Newtonian behavior. Their viscosity will increase dramatically above certain concentrations, making pumping difficult and reducing mass transfer within the boundary layer. This will eliminate most hollow fiber/capillary modules because they cannot withstand high pressure drops.

Membrane modules using flat sheets (spiral-wound, plate, and pleated sheet modules) usually have a meshlike spacer between sheets of membrane. This restricts their use to clear feed streams containing only fine SS.

Feed streams containing large SS would be treated poorly in spiral-wound modules, owing to the spacers in their feed channels. On the other hand, spiral-wound membrane modules are the lowest in capital costs and energy consumption. The trend in the food and beverage industries in recent years seems to be away from plate modules and towards spiral-wound modules, with ceramic tubular modules holding their own.

3. PROCESS COMPARISON

3.1. Similarity

Figure 4.6 and Table 4.1 show the similarities and dissimilarities of CAS process system and the innovated MBR process system.

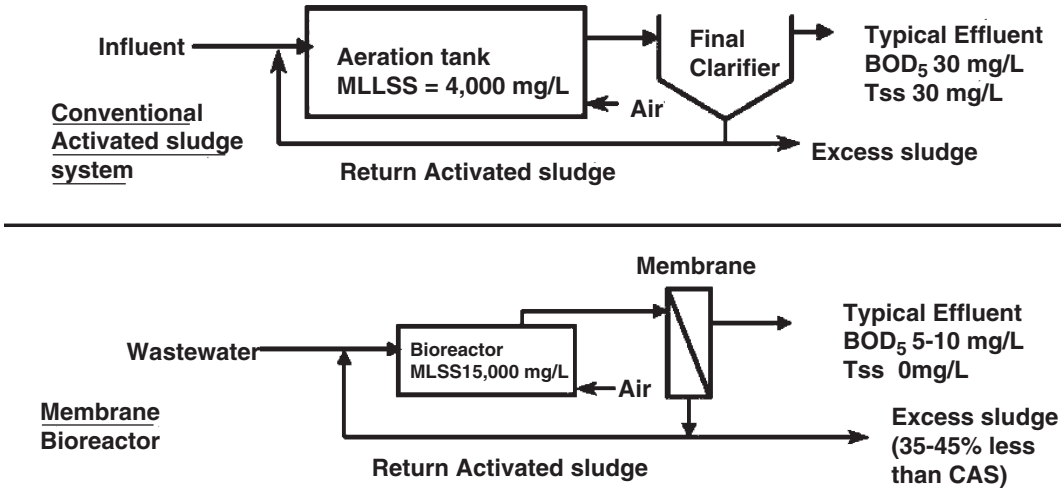

Fig. 4.6. Comparison between membrane bioreactor and conventional activated sludge system.

Table 4.1
Comparison of MBR and conventional activated sludge (CAS) systems

Dairy application	CAS	MBR
WW flow (m^3/day)	600	600
Influent COD (mg/L)	5,000	5,000
Influent BOD_5 (mg/L)	3,000	3,000
BOD_5 (kg/day)	1,800	1,800
Recycle of treated effluent (m^3/day)	0	400
Aeration volume (m^3)	4,500	600
Total floor space requirement (m^2)	1,300	260
Effluent COD (mg/L)	90	30
Effluent BOD_5 (mg/L)	30	5
Effluent TSS (mg/L)	30	0

Based on a 5-month pilot study at the dairy site in France.

The CAS and MBR process systems are similar from a biochemical engineering viewpoint. The basic process system of either CAS or MBR includes the unit processes of: influent feed, biological oxidation, final clarification, treated effluent discharge, return activated sludge (RAS), and excess sludge discharge. Both the CAS and MBR require air supply to support the biological oxidation. Because the theory and principles of process chemistry of both CAS and MBR are the same, the detailed process chemistry is discussed in the chapter on Activated Sludge Process, and will not be repeated here.

Both CAS and MBR can be operated for the purpose of carbonaceous oxidation, nitrification, and denitrification.

3.2. Dissimilarity

Then what are the differences between a CAS system and an MBR process system? Again the readers are referred to Fig. 4.6 and Table 4.1.

3.2.1. Reactor, MLSS, and Space Requirement Comparison

Assuming a CAS process system and a comparable MBR process system will be treating the same wastewater influent flow of 600 m^3/day, the influent BOD_5 concentration and BOD_5 load are 3,000 mg/L and 1,800 kg/day, respectively (*see* Table 4.1). The CAS system will require a large 4,500 m^3 aeration tank (MLSS = 4,000 mg/L) for biological oxidation, whereas the MBR system will require a much smaller 600 m^3 bioreactor (MLSS = 15,000 mg/L) for biological oxidation. As shown in Fig. 4.6, the mixed liquor from the CAS aeration tank flows to a large sedimentation clarifier (2 to 4 h hydraulic detention time) for clarification (or liquid–solid separation), whereas the mixed liquor from the MBR bioreactor flows to a compact membrane modules system (<0.5 h hydraulic detention time) for clarification. Accordingly, the footprint of an MBR process system is much smaller. In this specific example, the total floor space requirements of the CAS and MBR systems are 1,300 and 260 m^2, respectively.

3.2.2. Effluent Quality Comparison

The most significant comparison has been done on the treated effluent quality. Table 4.1 indicates that the effluent COD, BOD_5, and TSS are 90, 30, and 30 mg/L, respectively, for the CAS process system. The MBR process system's performance is much better: the effluent COD, BOD_5 and TSS are 30, 5, and 0 mg/L, respectively. The above process comparison is based on a 5-month pilot study at a dairy processing plant in France.

3.2.3. Cost Comparison and Water Recycle Considerations

How is the cost comparison? Based on limited cost data, the innovative MBR process system is cheaper to build, but more expensive to operate, in comparison with the CAS process system, if the treated effluent is not to be recycled for reuse.

Because of the high biomass concentration in the bioreactor (10,000 to 20,000 mg/L), the reactor can be made much more compact compared to CAS process systems. Additionally, this facilitates the system to accept higher organic loads.

Another major advantage of the MBR process system is that the excess sludge production is lower than conventional systems. This creates higher solids retention times (SRT) used in the process, and is a function of the shear forces imparted to the biomass as they move through the crossflow membrane units (for external membrane MBR configurations).

Thus, whereas the MBR system is an enhancement of the CAS system, it is quite different in space requirements, and especially in effluent quality. Figure 4.7 further illustrates graphically a comparison of the MBR and CAS systems, both designed to produce an effluent quality, suitable for recycle/reuse within and without the production facility.

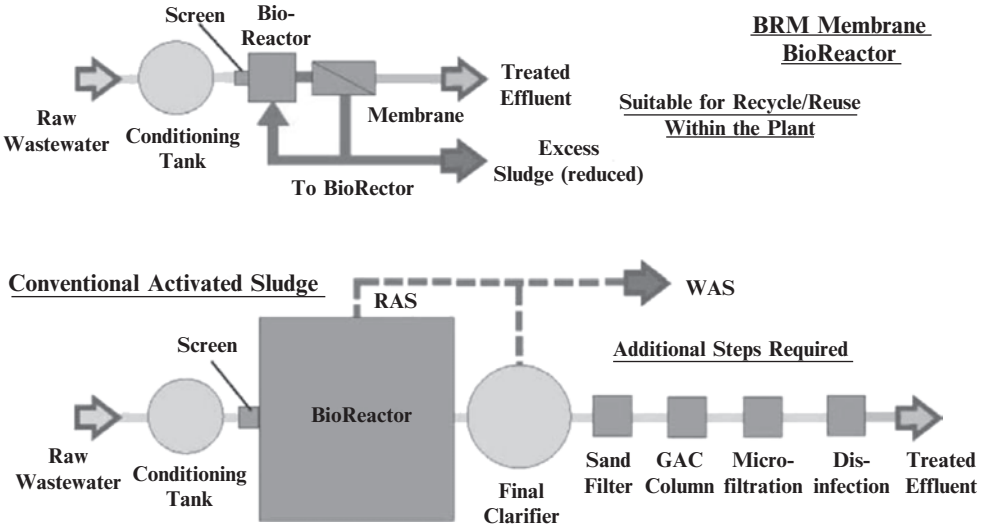

Fig. 4.7. Comparison between MBR and equivalent traditional WWTP.

In a water shortage region, such as California, the treated effluent should be recycled for reuse as much as possible. When water recycle is under consideration by environmental engineers, then both capital and O&M (Operations and Maintenance) costs of an MBR system will be much lower than that of a comparable CAS system. As shown in Table 4.1, 67% of the MBR treated effluent will meet the water quality requirements for direct nonpotable reuse, whereas the CAS treated effluent will not be suitable for recycle and reuse, unless tertiary treatment process units, such as, sand filter (SF), activated carbon filter (ACF), and disinfection are added for further effluent purification.

3.2.4. Waste Treatment Consideration

Finally the average MLSS in a CAS aeration tank is around 4,000 mg/L, whereas that for an MBR system is approximately 15,000 mg/L (10,000 to 20,000 mg/L range), as shown in Fig. 4.6. Then an MBR process system with much higher MLSS concentration is more suitable than a CAS process system when treating a high-strength wastewater stream.

3.2.5. Summary

In summation, the following are the advantages of an MBR process system over a CAS process system (15, 39, 40):

1. Excellent quality of treated effluent.
2. Possibility of recycle/reuse of treated effluent—better overall water economy.
3. Very compact installation: low construction costs.
4. Lower sludge production: lower sludge handling and nutrient costs.
5. Operating flexibility and simplicity; no sludge bulking problems, full automation possible.

6. Ideal preparation for the future; more stringent standards, rising costs of make-up water, etc.
7. Good aesthetics—appearance, odor, etc.
8. Modular design: easily expandable for future capacity.

4. PROCESS APPLICATIONS

4.1. Industrial Wastewater Treatment

The various advantages of the MBR process system give it a unique application niche in the treatment of industrial wastewater. Typical wastewater characteristics where MBR becomes a viable technology are as follows:

1. Flow rate: up to approximately 500,000 gpd.
2. COD: greater than approximately 2000 mg/L.

Industries where this technology can be implemented include chemical, petrochemical, pharmaceutical, fine chemicals, cosmetics, dairy, pulp and paper, automotive, landfill leachate, food, textiles, etc.

An MBR system has been designed for a petrochemical company located in south-east Texas to treat three high-strength industrial wastewaters containing alcohols and sulfur-containing compounds (17). The design was based on a field pilot test conducted by Envirogen, a company in Lawrenceville, New Jersey. One wastewater stream consisted of approximately 60% isopropanol by weight. The other streams contained light hydrocarbons and organic sulfides. The influent COD to the MBR system was 25,000 mg/L. Removal efficiencies averaged 90% to 95%, thereby allowing the plant to cost-effectively stay within regulatory limits. The three streams treated accounted for $<2\%$ of the plant's hydraulic wastewater load, but $>70\%$ of the organic wastewater load.

An industrial plant manager would like to consider possible adoption of an MBR process system for treating the industrial wastewater, usually because of the following reasons:

1. MBR system has smaller plant footprint because it treats low-flow high-strength streams, and operates at a much higher MLSS concentration.
2. MBR system has the possibility to recycle 40% treated effluent to existing RO step (no further pretreatment steps required).
3. MBR's modularity is suitable to double the capacity in the future.
4. MBR system can be installed easily in an old unused building.
5. MBR system has much less excess biosolids (sludge) production—its highly concentrated biosolids can be used as supplemental fuel in a boiler.
6. MBR system is the most cost-effective and reliable solution overall.
7. On-site pilot tests have shown simplicity and ease of operation of the MBR system.

4.2. Municipal Wastewater and Leachate Treatments

For treatment of high-flow low-strength municipal wastewater, the MBR process system can not economically compete with conventional activated sludge (CAS) process system, if

1. The municipality has plenty of land available for WWT facility construction

2. The treated effluent does not have to meet very stringent effluent standards (including nutrient removal and/or heavy metal removal)
3. The treated effluent does not have to be recycled for reuse
4. The project does not involve expansion of capacity or treatment by retrofit

In case one or more of the above factors does/do not apply, the MBR system will have an edge for competition with the CAS process system. The technical as well as economic feasibility of treating municipal wastewater by MBR has been positively demonstrated (17, 18).

For treatment of low-flow high-strength leachate from sanitary landfill sites, MBR process system is superior to the CAS process system in terms of both effluent quality and cost (15, 17, 20, 38).

Two aerobic MBR reactor systems are currently being designed for a municipal wastewater treatment district in southern New Jersey, USA. On of these MBR systems will be used to pretreat landfill leachate (17) shipped to the facility from the surrounding area. The effluent from the pretreatment system will then be polished in the existing municipal wastewater treatment plant (WWTP). The design influent flow to the MBR system is 400,000 gpd with a COD of 10,000 mg/L. COD is the measure of the amount of oxygen required to oxidize organic and oxidizable inorganic compounds in wastewater (WW). The COD test is used to determine the degree of pollution in WW. The footprint of the system is approximately 7000 ft^2 (2000 ft^2 for reactors and membranes and 5000 ft^2 for pumps, blowers, and other auxiliary equipment).

The second aerobic MBR system is being designed as a mobile publicly owned treatment works (POTW). It will be capable of treating 80,000 gpd with an influent BOD_5 of 625 mg/L. This system will have phosphorus removal and disinfection capabilities built in. The footprint for this system is approximately 640 ft^2. The system is trailer-mounted (two 40 ft long by 8 ft wide skids) and will be highway transportable. (17)

Another case history of an aerobic MBR system for the treatment of a sanitary landfill leachate (20) is presented in Section 5.2.

5. PRACTICAL EXAMPLES

5.1. Example 1. Dairy Industry

A dairy plant in central France produces 35,000 t/year of fruit- and other yogurts from milk. Its WW source includes washing of yogurt vat bottoms, other wash water, and cooling water blowdown. The plant needs a modern wastewater treatment (WWT) system to properly treat its combined WW. The following are the requirements:

1. Had very little floor space.
2. Water recycling (up to 70%) makes the plant less dependent on external water sources, which were not reliable.
3. The receiving water was in a fragile eco-system, which required as much flow and organic pollution to be removed.

Solution:

The modular design of the MBR process system enables the dairy plant to keep up with increased production in a phased manner, without over-investing in initial capital cost. Comparative data developed as a result of pilot testing are provided in Table 4.1. Both CAS and MBR process systems were piloted at the same time treating the same WW (Influent flow = 600 m^3/day, Influent COD = 5,000 mg/L, Influent BOD_5 = 3,000 mg/L) for 5 months.

It can be seen from Table 4.1 that the performance of the tested MBR process system was much better than that of the CAS process system.

Based on the pilot tests, the full-scale feasible CAS process system would have required 4,500 m^3 of aeration volume, and 1,300 m^2 of total floor area. Although the CAS system did meet the effluent COD standard (over 90% COD reduction), the effluent TSS averaged 30 mg/L, which was too high to recycle the CAS effluent for nonpotable reuse.

The MBR process system, on the other hand, was compact, requiring only 600 m^3 of aeration volume, and 260 m^2 of total floor area. The MBR effluent COD, BOD_5, and TSS were 30, 5, and 0 mg/L, respectively. The MBR effluent quality did meet the requirements for recycle as nonpotable water.

A full-scale MBR system was purchased by the dairy plant, and started up in May 1998, and is operating successfully.

The reasons for SLVO dairy plant to select MBR technology are summarized below:

1. High quality bacteria-free effluent (suitable for nonpotable reuse).
2. Possibility to recycle/reuse up to 70% of the treated effluent.
3. Small footprint (20% compared to CAS).
4. Ability to expand in the future.
5. Thirty-five percent less excess biosolids (waste sludge) production.
6. No odor problems.
7. Ease of operation and maintenance (operator friendly).
8. Fits in with equipment in the dairy plant.
9. Lowest cost option overall.

5.2. Example 2. Landfill Leachate Treatment

A landfill site in Arnouville, a small town in the suburbs of Paris, France needs a cost-effective process system for treating its municipal landfill leachate.

Landfill leachates originate mainly from percolation of rainwater and biological decomposition of wastes. Depending on such factors as age of the landfill and waste composition, leachates can contain high levels of organic and inorganic compounds, making treatment mandatory before reuse or discharge into the environment. Although conventional biological or physico-chemical processes (4–11) can efficiently remove SS, organic compounds and nitrogen, more stringent regulations have been implemented in several countries requiring removal of salts (chlorides, sulfates) and heavy metals.

Reverse osmosis (RO) is a well-known technology with many useful applications, mostly in the desalination of seawater. However, RO treatment of food industry process water (1)

Table 4.2
Performance of MBR–RO process system on landfill leachate application in Arnouville (near Paris), France

Parameters	Raw leachate quality	MBR effluent quality	RO effluent quality	Overall removal efficiency (%)
COD (mg/L)	2,500	710	10	>99
TOC (mg/L)	740	230	1	>99
NH_3-N (mg/L)	410	7	3	>99
Cl^- (mg/L)	1,500	1,450	50	>95
TSS (mg/L)	300	0	0	100

requires complete removal of SS and organic matter to avoid rapid fouling and clogging of the membranes (*see* Section 1.2.1).

In view of the respective capabilities of conventional biological processes, conventional physico-chemical processes, MBR, and RO, please recommend a solution to the landfill leachate treatment, which should be technically and economically feasible.

Solution:

In view of the respective capabilities of various processes, a combination of MBR processes and RO could provide an integrated system able to treat highly contaminated leachates and produce high-quality effluent meeting current and future regulations.

On the basis of results obtained during a 1-year pilot study, a full-scale plant was designed and installed to treat municipal leachates from a sanitary landfill site in the suburbs of Paris, France. The system consisted of an MBR process system followed by an RO unit. Results obtained are provided in Table 4.2.

Table 4.2 shows that the MBR process system (with ultrafiltration and/or microfiltration membrane) was able to achieve the following percent removal efficiency while treating the landfill leachate:

1. COD: 71.6%.
2. TOC: 68.9%.
3. NH_3-N: 98.3%.
4. Cl^-: 3.3%.
5. TSS: 100%.

Although the TSS removal for an MBR process system was 100%, the removals of COD, TOC, and NH_3-N were moderate-high, and that of chloride was poor.

The MBR system nevertheless was an excellent pretreatment unit for treating the sanitary landfill leachate, prior to the RO process system, due to reduction in silt density index (SDI), which is a very important parameter for satisfactory RO performance.

With the combination of MBR and RO, the overall removal efficiency of COD, TOC, NH_3-N, Cl^-, and TSS were all over 99%, which was very satisfactory.

The definitions of RO are given in Section 1.2.1.

Table 4.3
Coffee factory high-strength streams composition

	Flow		TSS		S-COD		T-COD		BOD_5	
Influent	gpd	gpm	mg/L	lb/day	mg/L	lb/day	mg/L	lb/day	mg/L	lb/day
Stream I	8,500	6	1,060	75	6,060	430	8870	630	3375	240
Stream II	24,000	17	220	44	6,440	1,290	7,000	1,400	3,375	675
Stream III	3,400	2	585	17	6,770	190	7,460	210	3,385	96
Stream IV	13,000	9	220	24	2,660	290	2,800	305	1,180	124
Total	48,900	34	390	160	5,390	2,200	6,240	2,545	2,780	1,135

Contaminant concentrations for the "Total" is an average based on "Total" load and flow.

5.3. Example 3. Coffee Industry

A coffee processing plant in Belgium produces 625 m^3/day (165,000 gpd) of combined WW, of which 70% is low-strength, and 30% is high-strength. The combined WW has the following characteristics:

1. BOD_5: 1,150 mg/L.
2. S-COD: 2,180 mg/L.
3. T-COD: 2,700 mg/L.
4. TSS: 280 mg/L.
5. Temperature: 15°C.

The high-strength WW was the plant manager's main concern. Table 4.3 summarizes the high-strength composition.

The government had issued an effluent discharge permit with the following effluent limitations:

1. Total Flow: 237,600 gpd.
2. TSS: 500 mg/L, and 220 lb/day.
3. S-COD: no limits.
4. T-COD: 2,000 mg/L and 3,000 lb/day.
5. BOD_5: 400 mg/L and 600 lb/day.

If you were the plant's environmental engineer responsible for WW compliance at this coffee plant, what would be your recommended engineering solutions to the plant manager?

Solution:

The coffee plant's environmental engineer decided to conduct a feasibility or treatability study, and selected an MBR pilot plant with the following specifications:

1. Skid dimensions = 13′ × 7′ × 8′H.
2. Weight = 4,000 lb (shipping); 8,000 lb (operating).
3. Connections =
   Influent = 1.5″ hose clamp

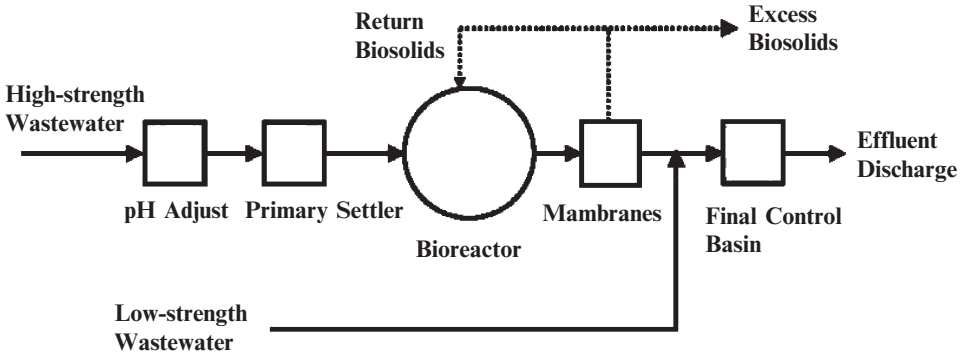

Fig. 4.8. Coffee factory WWT flow schematic.

Discharge = 2″ male NPT
Water supply = 5 to 10 gpm (3/4″ hose clamp)

4. Electrical = 3 phase, 240 V, 60 Hz, 100 A, 2 grounds.
5. Flow rate = approximately 1 gpm (depends on influent BOD/COD).

6. AUTOMATIC CONTROL SYSTEM

It was known that the MBR system is technically and economically feasible for treating high-strength and low-flow WW streams. It was then recommended by the plant's environmental engineer that only the low-flow high-strength stream (representing 30% of total combined WW flow) would require treatment in an MBR system. The remaining 70% untreated low-strength streams could be post-blended with the treated effluent from the MBR, resulting in a combined, or blended effluent, which would meet the discharge permit's effluent limitations (*see* Fig. 4.8).

The 1-gpm pilot plant demonstration was very successful. The above proposed engineering solution was fully proven by the MBR performance. Accordingly an MBR process system was ordered, installed, started-up, and operated at the coffee plant. Fig. 4.8 shows the full-scale WWT flow schematic implemented by the coffee plant. The successful performance data of the installed process system are shown in Table 4.4. It is seen from Table 4.4 that the quality of the MBR effluent was very high. Critical effluent parameters were as follows:

TSS = 0 mg/L
S-COD = 250 mg/L
T-COD = 250 mg/L
BOD_5 = 50 mg/L

After blending the MBR treated effluent and the untreated low-strength WW together, the resulting blended final effluent, indeed, met all effluent limitations in the permit.

Table 4.4
Coffee factory effluent characteristics

Effluent		MBR influent	MBR effluent	Untreated	Post-blended	Permitted
Flow	gpd	48,900	–	114,915	163,815	237,600
	gpm	34	–	80	114	165
TSS	mg/L	390	0	230	160	500
	lb/day	160	0	220	220	990
S-COD	mg/L	5,390	250	820	645	–
	lb/day	2,200	100	785	885	–
T-COD	mg/L	6,240	250	1,230	930	2,000
	lb/day	2,545	100	1,180	1,280	3,000
BOD_5	mg/L	2,780	50	460	335	400
	lb/day	1,135	20	440	460	600

6.1. Example 4. Cosmetics Industry

The WW discharged from a major cosmetics manufacturing plant in northern France was originally treated at the local municipal wastewater treatment plant (WWTP). Average flow rate was 160 m^3/day (42,240 gpd). In order to cut down sewer surcharge as well as fresh water costs, the plant set a goal to remove 90% of the total COD, and recycle at least 30% of the treated effluent for non-process uses within the plant.

Solution:

A pilot aerobic MBR test program was conducted at the plant using a 1 m^3 (264 US gal) pilot plant to determine treatability as well as to obtain full-scale design parameters. Results from the 5-month test program demonstrate the excellent overall performance of the MBR process system in terms of efficiency and reliability. Removal efficiencies obtained were 98 + % for COD, 99% for NH_4-N, and 99% for FOG (fats, oils, and greases). Removal of TSS was total; yielding an effluent that could satisfy the recycle criteria within the plant (15, 20).

Following the pilot test program, a full-scale system was designed and installed to handle 150 m^3/day flow, and 1200 kg/day COD. The membrane filtration unit consisted of ceramic microfiltration modules, which were modular and suitable for expansion. The plant has been successfully in operation since the summer of 1994. Despite the variable flow rate and characteristics of the influent WW (COD 2 to 6 g/L; COD/BOD_5 1.8 to 2.5), the treated effluent from the innovative MBR process system has been of consistent high quality (COD $<$ 100 mg/L; BOD $<$ 20 mg/L; TSS 0 mg/L; no bacteria). Part of the treated effluent is recycled for reuse in the factory lavatories, and for irrigation. The rest is discharged via a holding pond populated by fish, ducks, and reeds.

The quantity of excess biosolids produced is lower than conventional aerobic systems. Approximate net yield is 0.1 kg volatile suspended solids (VSS) per kg COD removed. The plant is easy and economical to operate. One part-time operator is adequate to perform process

Table 4.5
Aerobic MBR treatment of cosmetic factory wastewater in France

Parameter	Influent quality (mg/L)	Effluent quality (mg/L)
COD	6,500	<100
TSS	240	0
NH_3-N	40	0.4
FOG	2,500	<2

evaluation duties, which consist of standard analyses of WW parameters. An automatic system facilitates process control at this plant.

Table 4.5 presents the results of aerobic MBR treatment at this facility (20).

7. CONCLUSIONS

7.1. Industrial Applications

The MBR process system is a proven, reliable, modular, and compact industrial WWT system, that has been successfully implemented in several full-scale installations. Several years of operation has proven its reliability and efficiency for a variety of industrial plant owners, who use it as a water management tool, to conserve precious water resources and reduce overall operating costs (39, 40).

7.2. Municipal Applications

For treatment of high-flow low-strength WW, the MBR process system will be applicable and cost-effective if one or more of the factors below will apply:

1. The municipality has no space available for expansion.
2. The treated effluent must meet very stringent effluent standards (including nutrient and/or heavy metals removal).
3. The treated effluent has to be recycled for reuse.

In addition to the above municipal WWT applications, leachate from municipal sanitary landfills can also be cost-effectively treated by the newly developed MBR process system.

Although membrane process systems have been widely used for potable water treatment (3, 46), the use of MBR process systems for municipal potable water treatment (mainly aiming at nutrients removal) is still in developmental stage.

ACKNOWLEDGEMENT

The MBR process system was originally developed and perfected by Ondeo Degremont (Suez group), and Zenon Environmental Inc. Now the MBR process system has been accepted internationally as one of the mainstream biological WWT systems. More than 300 systems have been installed, and are successfully operated.

COMMERCIAL AVAILABILITY

Full-scale MBR process systems are available in the US through companies such as Ondeo Degremont, Zenon Environmental Inc., US Filter, etc.

REFERENCES

1. L. K. Wang and M. Cheryan, Application of Membrane Technology in Food Industry for Clean Production. *United Nationals Industrial Development Organization (UNIDO) Second International Conference on Waste Minimization and Cleaner Production, Barcelona, Spain, June 7–9, 1995. Technical Report No. DTT-8-6-95*, Vienna, Austria, 42p (1995).
2. L. K. Wang, J. V. Kropuzek, and U. Kounitson, Case Studies of Cleaner Production and Site Remediation. *UNIDO Technical Manual No. DTT-5-4-95, United Nationals Industrial Development Organization (UNIDO)*, Vienna, Austria, 136p (1995).
3. L. K. Wang, City of Cape Coral Reverse Osmosis Water Treatment Facility. *US Department of Commerce, National Technical Information Service, Technical Report No. PB97-139547*, Springfield, VA, 15p (1997).
4. L. K. Wang and N. C. Pereira, *Handbook of Environmental Engineering, Volume 3, Biological Treatment Processes*, p. 498, Humana Press, Totowa, NJ (1986).
5. C. Beer and L. K. Wang, Full-Scale operations of plug flow activated sludge systems. *Journal of New England Water Pollution Control Association*, 9, 2, 145–173 (1975).
6. L. K. Wang and M. H. S. Wang, Control tests and kinetics of activated sludge process. *Water, Air and Soil Pollution*, 8, 315–351 (1977).
7. C. Beer and L. K. Wang, Activated sludge systems using nitrate respiration—design considerations. *Journal of Water Pollution Control Federation*, 50, 9, 2120–2131 (1978).
8. M. Krofta and L. K. Wang, Improved biological treatment with a secondary flotation clarifier. *Civil Engineering for Practicing and Design Engineers*, 2, 307–324. (NTIS-PB82-154196) (1983).
9. L. K. Wang and D. B. Aulenbach, BOD and nutrient removal by biological A/O process systems. *US Departmen of Commerce, National Technical Information Service Technical Report #PB88-168430/AS*, Springfield, VA. 12p (1986).
10. L. K. Wang, Modern technologies for prevention and control of groundwater contamination. *Proceedings of New York–New Jersey Environmental Exposition*, NYNJEE, Belmont, MA (1990).
11. L. K. Wang and M. H. S. Wang, An overview of environmental biotechnologies with emphasis on aerobic and anaerobic fluidized bed. *Proceedings of Annual Convention of CAAPS*, Albany, NY (1990).
12. L. K. Wang, Wastewater treatment by biological-physicochemical two-stage process system. *Proceedings of the 41st Annual Purdue Industrial Waste Conference*, p. 67 (1987).
13. K. Brindle and T. Stephenson, The application of membrane biological reactors for the treatment of wastewaters. *Biotechnology and Bioengineering*, 49, 601 (1996).
14. N. Cicek, J. P. Franco, M. T. Suidan, and V. Urbain, Using a membrane bioreactor to reclaim wastewater. *Journal of American Water Works Association*, 90, 11, 105 (1998).
15. R. Menon and C. Fuchs, Application of degremont BRM^{TM} membrane bioreactor to the treatment of industrial wastewaters. *Membrane Technology Workshop at WEFTEC*, Orlando, FL (1998).
16. C. Chiemchaisri and K. Yamamoto, Biological Nutrient Removal under Low Temperatures in a Membrane Separation Bioreactor. *Water Science and Technology (G.B.)*, 28, 10, 325 (1993).
17. M. P. Pitre, D. N. Enegess, and R. Unterman, Bioreactors: the new wave in wastewater treatment. *Environmental Protection*, 10, 9, 30–33 (1999).

18. T. P. Giese, New wastewater treatment technologies. *Public Works*, 132, 5, 326–332 (2001).
19. Zenon Environmental Inc., Activated sludge without a clarifier. *Water Environment Federation*, 13, 9, 75 (2001).
20. A. Beaubien, E. Trouve, V. Urbain, D. Amar, and J. Manem, Membrane bioreactors offer new solution to old wastewater treatment problems. *Environmental Solutions*, Dec, 34–36 (1994).
21. V. Levy, The membrane bioreactor and the management of water resources in industry. *Presentation at POLLUTEC 96* (1996).
22. V. Urbain, R. Benoit, and J. Manem, Membrane bioreactor—a new treatment tool. *Journal of American Water Works Association*, 88, 75 (1996).
23. J. S. Kim, C. H. Lee, and H. D. Chun, Comparison of ultrafiltration characteristics between activated sludge and BAC sludge. *Water Research (G.B.)*, 32, 3443 (1998).
24. J. Dollorer and P. A. Wilderer, Biological treatment of leachates from hazardous waste landfills using SBBR technology. *Water Science Technology*, 34, 437 (1996).
25. A. G. Livingston, L. M. Freitas dos Santos, P. Pavasant, E. N. Pistikopoulos, and L. F. Strachen, Detoxification of industrial wastewaters in an extractive membrane bioreactor. *Water Science Technology*, 33, 1, 1 (1996).
26. L. F. Strachan, L. M. Freitas dos Santos, D. J. Leak, and A. G. Livingston, Minimization of biomass in an extractive membrane bioreactor. *Water Science Technology*, 34, 273 (1996).
27. P. Cote, H. Buisson, and M. Praderie, Immersed membranes activated sludge process applied to the treatment of municipal wastewater. *Water Science Technology (G.B.)*, 38, 437 (1998).
28. L. S. Chang, C. H. Lee, and K. H. Ahn, Membrane filtration characteristics in membrane-coupled activated sludge system—the effect of floc structure on membrane fouling. *Sep. Science and Technology*, 34, 9, 15 (1999).
29. J. A. Scott, J. A. Howell, T. C. Arnot, K. L. Smith, and M. Brusk, Enhanced system k_La and permeate flux with a ceramic membrane bioreactor. *Biotechnology Techniques*, 10, 4, 287 (1996).
30. M. G. Parratiyar, R. Govind, and D. F. Bishop, Treatment of trichloroethylene (TCE) in a membrane bioreactor. *Biotechnology and Bioengineering*, 50, 57 (1996).
31. L. Clapp, R. Hartono, M. Newman, and J. Park, Trichloroethylene degradation in a novel membrane bioreactor. *The 69th Annual Water Environment Federation Conference*, Dallas, TX (1996).
32. T. Nomura, T. Fuji, and M. Suzuki, Application of ceramic membrane with hydrophobic skin layer to separation of activated sludge. *Water Science and Technology (G.B.)*, 35, 4, 137 (1997).
33. K. H. Choo and C. H. Lee, Hydrodynamic behavior of anaerobic biosolids during crossflow filtration in the membrane anaerobic bioreactor. *Water Research (G.B.)*, 32, 3387 (1998).
34. L. van Dijk and G. C. G. Roncken, Membrane bioreactor for wastewater treatment: the state of the art and new developments. *Water Science and Technology (G.B.)*, 35, 10, 35 (1997).
35. B. Guender and K. Krauth, Replacement of secondary clarification by membrane separation—results with plate and hollow fiber modules. *Water Science and Technology (G.B.)*, 38, 383 (1998).
36. K. Yamamoto, M. Hiasa, T. Mahmood, and T. Matsuo, Direct solid-liquid separation using hollow-fiber membrane in an activated sludge aeration tank. *Water Science and Technology (G.B.)*, 21, 43 (1989).
37. K. H. Choo and H. D. Stensel, Sequencing batch membrane bioreactor treatment: nitrogen removal and membrane fouling evaluation. *Water Environment Research*, 72–4, 490–498.
38. D. Williams, Membrane treat wastewater and leachate. *Pollution Engineering*, 29, 1, 20–21 (1997).
39. J. Hensel and T. Mills, High and lows across the industry: 2002 executive forecast. *Environmental Protection*, 13, 1, 12–14 (2002).

40. S. Freeman, G. F. Leitner, J. Crook, and W. Vernon, A clear advantage. *Water Environment and Technology*, 14, 1, 16–21 (2002).
41. K. Abbott and H. Alpor, Protection from organic fouling. *Environmental Protection*, 12, 6, 34 (2001).
42. Y. Suwa, T. Suzuki, H. Toyohara, T. Yamagishi, and Y. Urushigawa, Single-stage nitrogen removal by an activated-sludge process with crossflow filtration. *Water Research (G.B.)*, 26, 1149 (1992).
43. M. C. M. van Loosdrecht and M. Henze, Maintenance, endogenous respiration, lysis, decay and predation. *Water Science and Technology (G.B.)*, 39, 1, 107 (1999).
44. US Filter, Immersed membrane bioreactor system. *Water Engineering and Management*, 150, 1, 4 (2003).
45. D. Pearson, Raisin producer's new process cuts wastewater and odors. *Water Engineering and Management*, 150, 1, 26–29 (2003).
46. R. Oreskovich, J. Contestables, K. Flat, I. C. Watson, and J. Rifleman, Reverse osmosis anion-filtration water plant. *Water Engineering and Management*, 150, 1, 10–19 (2003).
47. J. M. Hudkins and H. E. Schnidt, Wastewater Membrane Bioreactors—An Emerging Technology for the Development of a Reclaimed Water Supply. www.consulthai.com/2002.news/papers/waterReuse-2004-MBR (2004).
48. G. N. Sheindorf, N. Stahl, A. Tenenbaum, and Y. Levinsky, Membrane bioreactors for final treatment of wastewater. *Water Science Technology*, 48, 8, 103–110 (2003).
49. L. K. Wang and Z. Wu, Activated sludge processes. In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ (2009).
50. M. Glastra, H. V. Dijk, and J. Verberk, *Prospects of Membrane Bioreactors in Water Supply*. Berichte aus dem IWW Rheinisch-Westfalisches Institut fur Wasserforschung gemeinnutzige. Gmbtt Band 37b, Mulheim an der Rhur, 2002, ISSN 0941–0961 (2004).
51. L. K. Wang and Y. Li, Sequencing batch reactors, In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ (2009).
52. N. K. Shammas and L. K. Wang, Emerging suspended growth biological processes. In: *Advanced Biological Treatment Processes*, L. K. Wang, N. K. Shammas and Y. T. Hung (eds.), Humana Press, Totowa, NJ (2009).

5
SBR Systems for Biological Nutrient Removal

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract The sequencing batch reactor (SBR) is a fill-and-draw activated sludge system for wastewater treatment. SBR systems have been successfully used to treat both municipal and industrial wastewater. They are uniquely suited for wastewater treatment applications characterized by low or intermittent flow (IF) conditions. This chapter discusses the following aspects of the process: background and process description, proprietary SPR processes, description of a treatment plant using an SBR, applicability, advantages and disadvantages design criteria, process performance, operation and maintenance, cost, and packaged SBR for onsite systems.

Key Words Design criteria • O&M • cost • performance • SBR • Sequencing batch reactor.

1. BACKGROUND AND PROCESS DESCRIPTION

The sequencing batch reactor (SBR) is a fill-and-draw activated sludge system for wastewater treatment (1). The prototype for the activated sludge concept was developed on a

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_5

fill-and-draw basis (2). Shortly after that initial study, the emphasis switched to continuous flow "conventional" activated sludge. In an SBR system, wastewater is added to a single "batch" reactor, treated to remove undesirable components, and then discharged. Equalization, aeration, mixing, and clarification can all be achieved using a single batch reactor. To optimize the performance of the system, two or more batch reactors are used in a predetermined sequence of operations. SBR systems have been successfully used to treat both municipal and industrial wastewater. They are uniquely suited for wastewater treatment applications characterized by low or intermittent flow conditions.

Fill-and-draw batch processes similar to the SBR are not a recent development as commonly thought. Between 1914 and 1920, several full-scale fill-and-draw systems were in operation. Interest in SBRs was revived in the late 1950s and the early 1960s, with the development of new equipment and technology. Innovations in aeration devices, control logic, level sensors, solenoids, and hydraulic energy dissipators have surmounted the earlier limitations and revitalized interest in SBR technology (3). The resurgence of interest in SBRs was initially limited to small treatment applications; however, the need for greater treatment efficiencies and nutrient removal owing to increasingly stringent effluent limits has resulted in the adoption of SBR technology in installations as large as 660 L/s (15 MGD) (4).

The first modern, full-scale plant for SBR treatment of municipal wastewater in the United States was the Culver, Indiana, wastewater treatment facility (5). Retrofitted for the SBR process, operation was initiated in May 1980 (6). Since that time, SBR technology has become widespread in the United States, with more than 150 plants in operation (7). SBRs can be modified to provide carbonaceous oxidation, nitrification, and biological nutrient removal (BNR). Approximately 25% of all SBR systems were designed to achieve nutrient removal (8).

The unit processes of the SBR and conventional activated sludge systems are the same. A US EPA report summarized this by stating that the SBR is no more than an activated sludge system that operates in time rather than in space (1, 3). The difference between the two technologies is that the SBR performs equalization, biological treatment, and secondary clarification in a single tank using a timed control sequence. This type of reactor does, in some cases, also perform primary clarification. In a conventional activated sludge system, these unit processes would be accomplished by using separate tanks.

The SBR consists of a self-contained treatment system incorporating equalization, aeration, anoxic reaction, and clarification within one basin. Intermittently fed SBRs consist of the following basic steps (1, 3, 9):

1. ***Fill***—The fill operation consists of adding the waste and substrate for microbial activity. The fill cycle can be controlled by float switches to a designated volume or by timers for multireactor systems. A simple and commonly applied mode to control the fill cycle is based on reactor volume, resulting in fill times inversely related to influent flow rates. The fill phase can include many phases of operation and is subject to various modes of control, termed static fill, mixed fill, and react fill. Static fill involves the introduction of waste influent with no mixing or aeration. This type of fill method is most common in plants requiring nutrient control. In such applications, the static fill will be accompanied by a mixed fill stage such that the microorganisms are exposed to sufficient substrate, while maintaining anoxic or anaerobic conditions. Both mixing and aeration

are provided in the react fill stage. The system may alternate among static fill, mixed fill, and react fill throughout the fill cycle.

2. ***React***—The purpose of the react stage is to complete reactions initiated during fill. The react stage may be comprised of mixing or aeration, or both. As was the case in the fill cycle, desired processes may require alternating cycles of aeration, The length of the react phase may be controlled by timers, by liquid level controls in a multitank system, or when the desired degree of treatment has been attained, verified by monitoring of reactor contents. Depending upon the amount and timing of aeration during fill, there may or may not be a dedicated react phase.
3. ***Settle***—Liquid–solid separation occurs during the settle phase, analogous to the operation of a conventional final clarifier. Settling in an SBR can demonstrate higher efficiencies than a continuous-flow settler, since total quiescence is achieved in an SBR.
4. ***Draw***—Clarified effluent is decanted in the draw phase. Decanting can be achieved by various apparatus, the most common being floating or adjustable weirs. The decanting capability is one of the operational and equipment limitations of SBR technology. Adaptation or development of equipment compatible with a fluctuating liquid level is required.
5. ***Idle***—The final phase is termed the idle phase and is only used in multibasin applications. The time spent in the idle phase will depend on the time required for the preceding basin to complete its fill cycle. Biosolids wastage will typically be performed during the idle phase.

A typical SBR process sequence schematic is shown in Fig. 5.1.

Denitrification can occur during the fill or react stages by cycling the aerators, and during the settle and draw period. An obvious advantage of an SBR systems with low flows is that the reactor contents can be retained until the desired level of treatment is achieved, providing that sufficient tankage exists to equalize or accommodate the additional influent.

2. PROPRIETARY SBR PROCESSES

SBR manufacturers have adapted the sequence of batch treatment cycles in various ways. One classification of SBR systems distinguishes those which operate with continuous feed and intermittent discharge (CFID) from those which operate with intermittent feed and intermittent discharge (IFID). IFID reactors are characteristic of the conventional fill-and-draw SBR reactors in that the influent flow to the reactor is discontinued for some portion of each cycle. The CFID reactors receive wastewater during all phases of the treatment cycle. A key design consideration with such systems is minimization of short-circuiting between influent and effluent. This is accomplished by locating the feed and withdrawal points at opposite ends of the tank, using rectangular reactors with length-to-width ratios of at least 2 to 1 and providing baffling.

The steps and associated conditions and purpose of a complete, typical cycle for a single tank operated as part of an IFID SBR system designed to achieve nitrification are described in Table 5.1. Nitrification takes place during the react phase and during the portions of the fill period when aeration is practiced.

Several proprietary process and equipment innovations have been developed to enhance treatment, simplify operation, or control biosolids characteristics (9–15). All proprietary SBR manufacturers will guarantee TN effluent concentrations <5 mg/L. To illustrate the variety of options available, the proprietary aspects of five SBR manufacturers are discussed below.

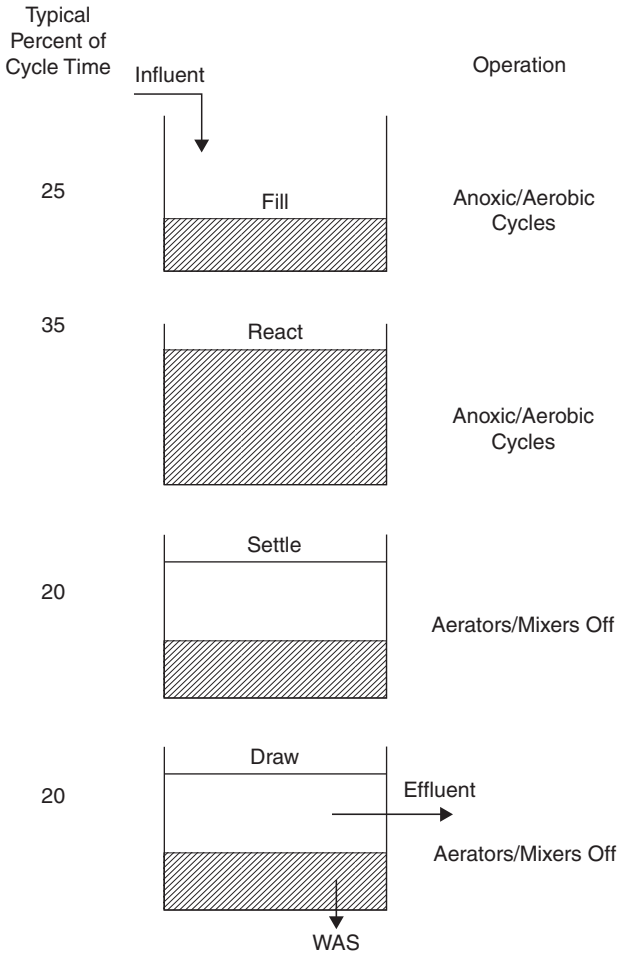

Fig. 5.1. Sequencing batch reactor (SBR). (*Source*: US EPA).

2.1. Aqua SBR

The Aqua SBR system provided by Aqua-Aerobic Systems, Inc. (11) is not a patented process, but the process does include a proprietary floating direct-drive mixer, an effluent decanter, and a microprocessor control system. The floating decanter is designed to prohibit MLSS from entering the decanter during mixed or react phases, and it also withdraws supernate 30 cm (0.5 ft) below the water surface to mitigate scum losses to the effluent. If long settling times are provided, clear effluent can be obtained at high sludge volume index (SVI).

Table 5.1
Typical cycle for a single tank in a dual tank SBR system designed for nitrification

Step	Conditions	Purpose
FILL	Influent flow into SBR Aeration occurs continually or intermittently Time = half of cycle time	Addition of raw wastewater to the SBR; COD removal and nitrification
REACT	No influent flow to SBR Aeration Time typically = 1 to 2 h (varies widely depending on nitrification kinetics, waste strength, and amount of aeration during fill)	Carbonaceous oxidation and nitrification
SETTLE	No influent flow to SBR No aeration Time = approximately 1 h (depends on settling characteristics)	Allow SS to settle, yielding a clear supernatant
DRAW	No influent flow to SBR No aeration Effluent is decanted Time = 1 h (variable)	Decant—remove clarified effluent from reactor; 15% to 25% of the reactor volume is typically decanted, depending on hydraulic considerations and SBR manufacturer's design
IDLE	No influent flow to SBR No aeration Sludge is wasted Time = variable (determined by flow rate)	Multitank system, which allows time for one reactor to complete the fill step before another starts a new cycle; waste sludge—remove excess solids from reactors

Source: US EPA.
A typical total cycle time is 4 to 6 h.

2.2. Omniflo

Jet Tech, Inc. (12) has developed SBR equipment and also has a patented logic control for their aeration system. The proprietary equipment includes dry pit pumps, headers, manifolds, influent distribution hardware, jet aerators, and decanter apparatus. A proprietary aspect of the SBR process provided by Jet Tech is the Batch Proportional Aeration System. The function of this aeration system is to relate the volumetric change rate during the fill phase to the aeration capacity requirements by sensing the DO level in the reactor, optimizing nitrification and denitrification cycles.

2.3. Fluidyne

The Fluidyne Corp. (13, 14) offers a system with effluent decanters fixed in position to the reactor wall. The device excludes missed liquor suspended solids (MLSS) entry during aeration. These systems also commonly employ jet aeration with a combination of aeration and static conditions during fill.

2.4. CASS

The cyclic activated sludge system (CASS) was developed and is marketed by Transenviro, Inc. CASS uses a similar sequence of operation as other batch technologies, but is configured with a proprietary captive selector reactor. The selector can also receive continuous flow. The selector is a baffled compartment that receives raw wastewater or primary effluent where it is mixed with RAS or internally recycled MLSS. The selector then conveys flow to the reactor basin. By limiting or eliminating aeration to the selector, oxygen deficient conditions can be attained, while concurrent high substrate levels are maintained. This mode of operation is claimed to favor the propagation of floc formers and to inhibit growth of filamentous strains (15). A process schematic is presented in Fig. 5.2.

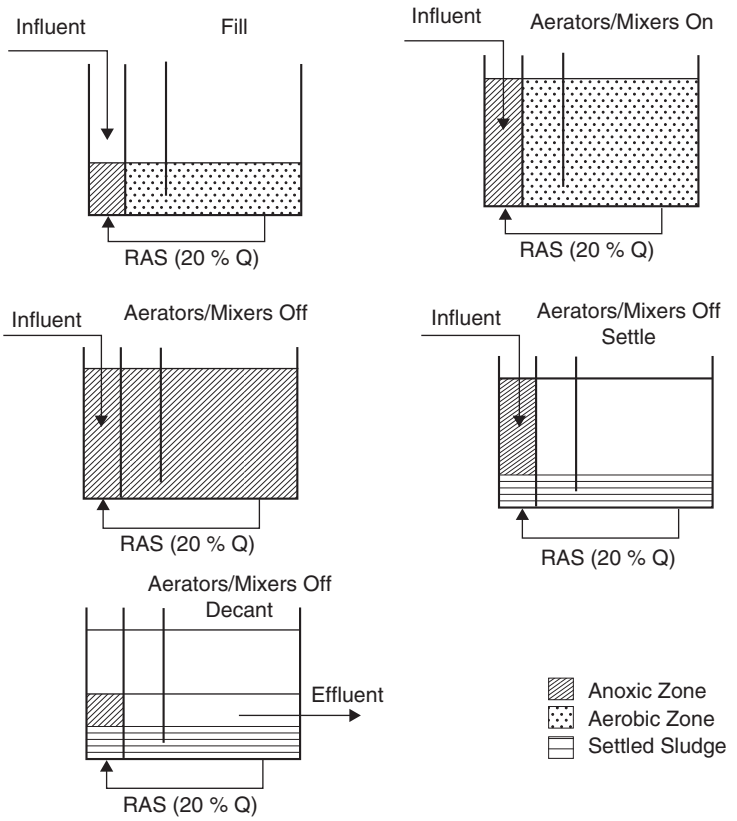

Fig. 5.2. Cyclical activated sludge system (CASS). (*Source*: US EPA).

2.5. ICEAS

A modified batch system is available from Austgen-Biojet (ABJ). The ABJ system is termed intermittent cycle extended aeration system (ICEAS) and is depicted schematically in Fig. 5.3. The distinguishing feature of ICEAS is that continuous inflow is incorporated in all phases, compared to other variable volume processes that do not receive continuous inflow. Noncontinuous inflow operation can be provided, if requested. Austgen-Biojet maintains that the continuous inflow mode is preferable to noncontinuous flow operation, as the distribution box used by ABJ will ensure that variations in load and flow are distributed evenly between the reactors and prevent diurnal variations or shock loads from continually overloading one reactor. The manufacturer asserts an additional advantage of the ICEAS flow regime is that continuous flow via the distribution box reduces the valving and headworks engineering compared to requirements for a noncontinuous flow SBR. A complete ICEAS treatment cycle consists of three phases: aeration, settle, and draw. Because influent is received during all phases, ICEAS does not offer total quiescence during the settle phase,

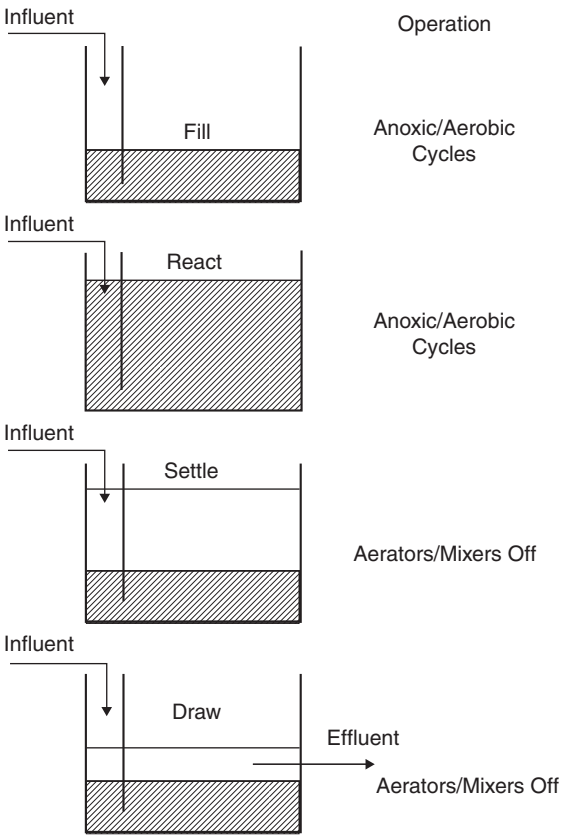

Fig. 5.3. Intermittent cycle extended aeration system. (*Source*: US EPA).

a characteristic of an intermittently fed SBR. Although ICEAS is proprietary, no royalty or license fees are imposed. ICEAS uses a patented anoxic selector to provide denitrification and to promote growth of zoogleal microorganisms, and to inhibit filamentous strains. The ABJ selector has characteristics similar to the patented CASS selector, but ABJ claims to be the developer of the original selector concept.

3. DESCRIPTION OF A TREATMENT PLANT USING SBR

A typical process flow schematic for a municipal wastewater treatment plant using an SBR is shown in Fig. 5.4 (1, 3). Influent wastewater generally passes through screens and grit removal before the SBR. The wastewater then enters a partially filled reactor, containing biomass, which is acclimated to the wastewater constituents during preceding cycles. Once the reactor is full, it behaves like a conventional activated sludge system, but without a continuous influent or effluent flow. The aeration and mixing is discontinued after the biological reactions are complete, the biomass settles, and the treated supernatant is removed. Excess biomass is wasted at any time during the cycle. Frequent wasting results in holding the mass ratio of influent substrate to biomass nearly constant from cycle to cycle. Continuous flow systems hold the mass ratio of influent substrate to biomass constant by adjusting return activated sludge (RAS) flowrates continually as influent flowrates, characteristics, and settling tank underflow concentrations vary. After the SBR, the "batch" of wastewater may flow to an equalization basin where the wastewater flow to an additional processing unit can be controlled at a determined rate. In some cases the wastewater is filtered to remove additional solids and then disinfected.

As illustrated in Fig. 5.4, the solids handling system may consist of a thickener and an aerobic digester. With SBRs there is no need for return activated sludge (RAS) pumps and primary sludge (PS) pumps like those associated with conventional activated sludge systems. With the SBR, there is only one sludge biomass (biosolids) to handle. The need for gravity thickeners before digestion is determined on a case by case basis depending on the characteristics of the biosolids.

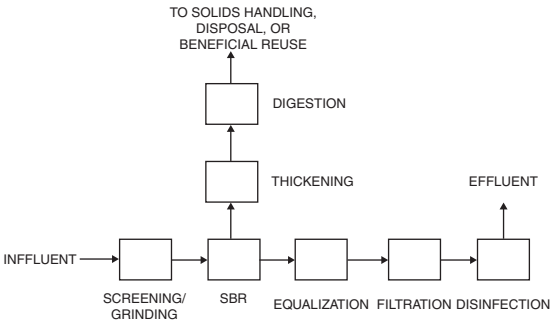

Fig. 5.4. SBR process flow diagram. (*Source*: US EPA).

An SBR serves as an equalization basin when the vessel is filling with wastewater, enabling the system to tolerate peak flows or peak loads in the influent and to equalize them in the batch reactor. In many conventional activated sludge systems, separate equalization is needed to protect the biological system from peak flows, which may wash out the biomass, or peak loads, which may upset the treatment process.

It should also be noted that primary clarifiers are typically not required for municipal wastewater applications before an SBR. In most conventional activated sludge wastewater treatment plants, primary clarifiers are used before the biological system. However, primary clarifiers may be recommended by the SBR manufacturer if the total suspended solids (TSS) or biochemical oxygen demand (BOD) are greater than 400 to 500 mg/L. Historic data should be evaluated and the SBR manufacturer consulted to determine whether primary clarifiers or equalization are recommended before an SBR for municipal and industrial applications.

Equalization may be required after the SBR, depending on the downstream process. If equalization is not used before filtration, the filters need to be sized to receive the batch of wastewater from the SBR, resulting in a large surface area required for filtration, Sizing filters to accept these "batch" flows is usually not feasible, which is why equalization is used between an SBR and downstream filtration. Separate equalization following the biological system is generally not required for most conventional activated sludge systems, because the flow is on a continuous and more constant basis.

4. APPLICABILITY

SBRs are typically used at flowrates of 5 MGD or less (1, 3). The more sophisticated operation required at larger SBR plants tends to discourage the use of these plants for large flowrates. The SBR technology is particularly attractive for treating smaller wastewater flows. The majority of plants were designed at wastewater flow rates of less than 22 L/s (0.5 MGD) (7). The cost-effectiveness of SBRs may limit their use to flows less than 440 L/s (10 MGD) (6). Depending on the number of SBR reactors in a plant and the duration of the discharge cycle, the downstream units often must be sized for two or more times the influent flow rate. Plants with four or more separate reactors may have the reactor process cycles offset such that the discharge is nearly continuous.

As these systems have a relatively small footprint, they are useful for areas where the available land is limited. In addition, cycles within the system can be easily modified for nutrient removal in the future, if it becomes necessary. This makes SBRs extremely flexible to adapt to regulatory changes for effluent parameters such as nutrient removal. SBRs are also very cost effective if treatment beyond biological treatment is required, such as filtration.

5. ADVANTAGES AND DISADVANTAGES

Some advantages and disadvantages of SBRs are listed below (1, 3, 8):

Advantages

1. Equalization and the ability to tolerate peak flows and shock loads of BOD_5.
2. Primary clarification (in most cases), biological treatment, and secondary clarification can be achieved in a single reactor vessel.

3. Operating flexibility and control of effluent discharge.
4. Minimal footprint.
5. Potential capital cost savings by eliminating clarifiers and other equipment.

Disadvantages

1. A higher level of sophistication is required (compared to conventional systems), especially for larger systems, of timing units and controls.
2. Higher level of maintenance (compared to conventional systems) associated with more sophisticated controls, automated switches, and automated valves.
3. Potential of discharging floating or settled biosolids during the draw or decant phase with some SBR configurations.
4. Potential plugging of aeration devices during selected operating cycles, depending on the aeration system used by the manufacturer.
5. Potential requirement for equalization after the SBR, depending on the downstream processes.

6. DESIGN CRITERIA

For any wastewater treatment plant design, the first step is to determine the anticipated influent characteristics of the wastewater and the effluent requirements for the proposed system. These influent parameters typically include design flow, maximum daily flow BOD_5, TSS, pH, alkalinity, wastewater temperature, total Kjeldahl nitrogen (TKN), ammonia-nitrogen (NH_3-N), and total phosphorus (TP). For industrial and domestic wastewater, other site specific parameters may also be required.

The state regulatory agency should be contacted to determine the effluent requirements of the proposed plant. These effluent discharge parameters will be dictated by the state in the National Pollutant Discharge Elimination System (NPDES) permit. The parameters typically permitted for municipal systems are flowrate, BOD_5, TSS, and fecal coliform (FC). In addition, many states are moving toward requiring nutrient removal. Therefore, total nitrogen (TN), TKN, NH_3-N, or TP may also be required. It is imperative to establish effluent requirements because they will impact the operating sequence of the SBR. For example, if there is a nutrient requirement and NH_3-N or TKN is required, then nitrification will be necessary. If there is a TN limit, then nitrification and denitrification will be necessary.

6.1. Design Parameters

Once the influent and effluent characteristics of the system are determined, the engineer will typically consult SBR manufacturers for a recommended design. Based on these parameters, and other site specific parameters such as temperature, key design parameters are selected for the system. An example of these parameters for a wastewater system loading is listed in Table 5.2.

A unified approach to SBR technology has yet to be developed (16); however, the principles used to design nitrification–denitrification facilities in single anoxic or dual anoxic zone systems, such as flow and loadings, may be applied with some modifications. One factor to consider specifically for the design of an SBR is the flow volume that will determine whether one reactor will suffice (generally for flows <2 L/s or 0.05 MGD) or whether a

Table 5.2
SBR design parameters for conventional load

	Municipal	Industrial
Food to Mass (F/M)	0.15-0.4/day	0.15-0.6/day
Treatment cycle duration	4.0 h	4.0-24 h
Typically low water level mixed liquor suspended solids	2000-2500 mg/L	2000-4000 mg/L
Hydraulic retention time	6-14 h	Varies

Source: US EPA.

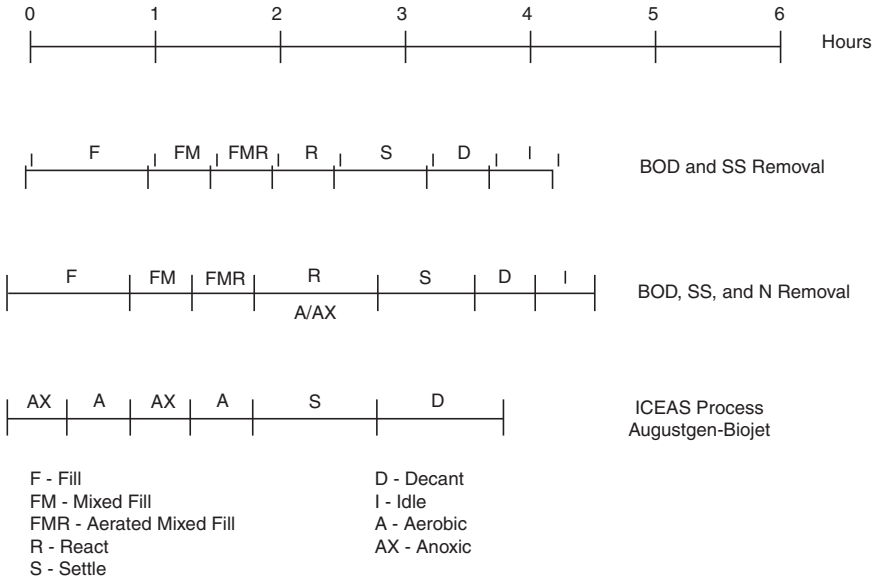

Fig. 5.5. Operating strategies for SBR systems. (*Source*: US EPA).

two-vessel system is required. Additional vessels should be considered for sites that experience a wide transient variation in either organic or hydraulic loading. Conditions, including wet weather with ingress of surface or ground waters, may be accommodated by effecting more frequent decant cycles, without causing washout of the reactor biomass. The SBR process can accommodate peak hourly flows three to ten times as large as the design flow without adverse effects, if excess capacity is available. The F/M ratio must be determined by the desired effluent quality which in turn dictates reactor sizing.

The critical operational feature is the cycle time for fill, react, settle, and draw, and the amount of oxygen that is supplied. A typical cycle for an intermittent-feed, intermittent-discharge SBR based on average flow conditions is 4-hour duration; 2 hour allocated to fill/aeration/anoxic react, 1 hour to settling, and 1 hour to decant and idle. The total time for a batch cycle consists of the time allowed for each component phase. Design cycle times in

Table 5.3
Typical design criteria for SBRs

Parameter	SBR	ICEAS
BOD load ($g/d/m^3$)	80–240	
Cycle time (h)		
Fill (aeration)	1–3	
Settle	0.7–1	
Draw	0.5–1.5	
MLSS (mg/L)	2300–5000	
MLVSS (mg/L)	1500–3500	
HRT (h)	15–40	36–50
θ_c (day)	20–40	–
F/M (g BOD_5/g MLVSS/day)	0.05–0.20	0.04–0.06

Source: US EPA.

full-scale plants have varied from 2 to 24 h (17). A suggested strategy is presented in Fig. 5.5. Some typical design criteria are presented in Table 5.3.

SBR systems are typically designed and operated at long solids residence times (> 15 days) and low F/M (less than 0.1 kg BOD_5/kg MLSS/day). Consequently, partial or complete nitrification is nearly always observed (7, 8). In an evaluation of 19 SBR treatment plants (8) (all originally designed for nitrification), influent and effluent ammonia-nitrogen data were reported for eight of the plants (Table 5.4). The average effluent ammonium-nitrogen concentration for the eight plants was less than 2.0 mg/L, implying that a high degree of nitrification was achieved in all cases. These efficiencies reflect the long design solids residence times that are employed and operations that are generally well below the design flow.

The design mixed liquor volume can be calculated from the selected MLSS concentration, which decreases throughout the fill cycle. The MLSS concentration at the end of the draw phase is that of settled mixed liquor and is similar to that in a conventional clarifier underflow (18). Once the tank volumes have been calculated, the cycle times can be determined. If the cycle times are unsatisfactory, the tank volumes can be adjusted accordingly. The number of cycles per day, number of basins, decants volume, reactor size, and detention times can then be calculated.

Other site-specific information is needed to size the aeration equipment, such as site elevation above mean sea level, wastewater temperature, and total dissolved solids concentration. The sizing of aeration equipment is done according to criteria for complete nitrification and BOD removal, except that the required oxygen transfer must be accomplished in a shorter period. The actual amount of aeration time per cycle must be considered when sizing the aeration equipment.

The operation of an SBR is based on the fill-and-draw principle, which, as discussed in a previous section, consists of five basic steps: idle, fill, react, settle, and draw. More than one operating strategy is possible during most of these steps. For industrial wastewater applications, treatability studies are typically required to determine the optimum operating

Table 5.4
Nitrification performance information for SBR operating plants

Plant location	Period of evaluation	Wastewater flow		Percent of design flow	BOD_5 (mg/L)		Ammonia-N (mg/L)	
		m^3/day	MGD		Influent	Effluent	Influent	Effluent
Buckingham, PA	04/89–04/91	439	0.116	49	324	8	25.3	1.1
Clarkston, MI (Chateau Estates)	11/89–04/91	208	0.055	50	192	12	39.1	1.7
Grundy Center, IA	12/89–11/90	2176	0.575	72	195	4	15.8	1.2
Marlette, MI	07/90–06/91	1578	0.417	60	103	4	10.1	0.5
Mifflinburg, PA	10/88–03/91	2763	0.73	81	105	12	7.8	0.4
Monticello, IN (White Oaks Resort)	10/89–05/91	15	0.004	8	131	5	3.1	0.3
Muskegon Heights, MI (Clover Estates)	01/88–10/90	132	0.035	78	185	9	21.2	0.7
Windgap, PA	02/90–10/90	2116	0.559	56	160	7	12.9	0.6

Source: US EPA.
[a] Average monthly values based on all data available.

sequence. For most municipal wastewater treatment plants, treatability studies are not required to determine the operating sequence because municipal wastewater flowrates and characteristic variations are usually predictable and most municipal designers will follow conservative design approaches.

The idle step occurs between the draw and the fill steps, during which treated effluent is removed and influent wastewater is added. The length of the idle step varies depending on the influent flowrate and the operating strategy. Equalization is achieved during this step if variable idle times are used. Mixing to condition the biomass and biosolids wasting can also be performed during the idle step, depending on the operating strategy.

Influent wastewater is added to the reactor during the fill step. The following three variations are used for the fill step and any or all of them may be used depending on the operating strategy: static fill, mixed fill, and aerated fill. During static fill, influent wastewater is added to the biomass already present in the SBR. Static fill is characterized by no mixing or aeration, meaning that there will be a high substrate (food) concentration when mixing begins. A high food to microorganisms (F/M) ratio creates an environment favorable to floc forming organisms versus filamentous organisms, which provides good settling characteristics for the biosolids. Additionally, static fill conditions favor organisms that produce internal storage products during high substrate conditions, a requirement for biological phosphorus removal. Static fill may be compared to using "selector" compartments in a conventional activated sludge system to control the F/M ratio.

Mixed fill is classified by mixing influent organics with the biomass, which initiates biological reactions. During mixed fill, bacteria biologically degrade the organics and use residual oxygen or alternative electron acceptors, such as nitrate-nitrogen. In this environment, denitrification may occur under these anoxic conditions. Denitrification is the biological conversion of nitrate-nitrogen to nitrogen gas. An anoxic condition is defined as an environment in which oxygen is not present and nitrate-nitrogen is used by the micro-organisms as the electron acceptor. In a conventional biological nutrient removal (BNR) activated sludge system, mixed fill is comparable to the anoxic zone that is used for denitrification. Anaerobic conditions can also be achieved during the mixed fill phase. After the micro-organisms use the nitrate-nitrogen, sulfate becomes the electron acceptor. Anaerobic conditions are characterized by the lack of oxygen and sulfate as the electron acceptor.

Aerated fill is classified by aerating the contents of the reactor to begin the aerobic reactions completed in the react step. Aerated fill can reduce the aeration time required in the react step.

The biological reactions are completed in the react step, in which mixed react and aerated react modes are available. During aerated react, the aerobic reactions initialized during aerated fill are completed and nitrification can be achieved. Nitrification is the conversion of ammonia-nitrogen to nitrite-nitrogen and ultimately to nitrate-nitrogen. If the mixed react mode is selected, anoxic conditions can be attained to achieve denitrification. Anaerobic conditions can also be achieved in the mixed react mode for phosphorus removal.

Settle is typically provided under quiescent conditions in the SBR. In some cases, gentle mixing during the initial stages of settling may result in a clearer effluent and a more concentrated settled biosolids. In an SBR, there are no influent or effluent currents to interfere with the settling process as in a conventional activated sludge system.

The draw step uses a decanter to remove the treated effluent, which is the primary distinguishing factor between different SBR manufacturers. In general, there are floating decanters and fixed decanters. Floating decanters offer several advantages over fixed decanters as described in the Tank and Equipment Description section.

SBR technology requires unique and innovative strategies to accomplish each phase of the process cycle. Large facilities that require dual vessels can accommodate continuous flow by alternating fill cycles between reactors; single-vessel facilities except for ICEAS systems will require flow equalization or a selector. Compartments or baffles may be included within a selector to control the hydraulic regime and biosolids characteristics. Several criteria have been proposed that can be used to design an appropriate selector (19, 20). The CASS process by Transenviro is a proprietary SBR that includes an integral selector as part of the process. For more details on SBR design the readers are referred to Wilderer et al (21) and Toby (22).

6.2. Construction

Construction of SBR systems can typically require a smaller footprint than conventional activated sludge systems because the SBR often eliminates the need for primary clarifiers. The SBR never requires secondary clarifiers. The size of the SBR tanks themselves will be site specific; however the SBR system is advantageous if space is limited at the proposed site. A few case studies are presented in Table 5.5 to provide general sizing estimates at different flowrates. Sizing of these systems is site specific and these case studies do not reflect every system at that size.

SBR reactors have been constructed with a variety of shapes including rectangular, oval, circular, and with sloped sidewalls. Design bottom water levels after decant are typically 3 to 4 m (10 to 13 ft) and design top water levels are typically 4.3 to 5.5 m (14 to 18 ft). A freeboard of 1 m (3 ft) is common.

Table 5.5
Case studies for several SBRs facilities

Flow	Reactors			Blowers	
(MGD)	No.	Size (ft)	Volume (MG)	No.	Size (HP)
0.012	1	18 × 12	0.021	1	15
0.10	2	24 × 24	0.069	3	7.5
1.2	2	80 × 80	0.908	3	125
1.0	2	58 × 58	0.479	3	40
1.4	2	69 × 69	0.678	3	60
1.46	2	78 × 78	0.910	4	40
2.0	2	82 × 82	0.958	3	75
4.25	4	104 × 80	1.556	5	200
5.2	4	87 × 87	1.359	5	125

Source: US EPA.

The actual construction of the SBR tank and equipment may be comparable or simpler than a conventional activated sludge system. For Biological Nutrient Removal (BNR) plants, an SBR eliminates the need for return activated sludge (RAS) pumps and pipes. It may also eliminate the need for internal mixed liquor suspended solid (MLSS) recirculation, if this is being used in a conventional BNR system to return nitrate-nitrogen.

The control system of an SBR operation is more complex than a conventional activated sludge system and includes automatic switches, automatic valves, and instrumentation. These controls are very sophisticated in larger systems. The SBR manufacturers indicate that most SBR installations in the United States are used for smaller wastewater systems of less than 2 MGD and some references recommend SBRs only for small communities where land is limited. This is not always the case, however, as the largest SBR in the world is currently a 10 MGD system in the United Arab Emirates (23).

6.3. Tank and Equipment Description

The SBR system consists of a tank, aeration, and mixing equipment, a decanter, and a control system. The central features of the SBR system include the control unit and the automatic switches and valves that sequence and time the different operations. SBR manufacturers should be consulted for recommendations on tanks and equipment. It is typical to use a complete SBR system recommended and supplied by a single SBR manufacturer. It is possible, however, for an engineer to design an SBR system, as all required tanks, equipment, and controls are available through different manufacturers. This is not typical of SBR installation because of the level of sophistication of the instrumentation and controls associated with these systems.

The SBR tank is typically constructed with steel or concrete. For industrial applications, steel tanks coated for corrosion control are most common whereas concrete tanks are the most common for municipal treatment of domestic wastewater. For mixing and aeration, jet aeration systems are typical as they allow mixing either with or without aeration, but other aeration and mixing systems are also used. Positive displacement blowers are typically used for SBR design to handle wastewater level variations in the reactor. The varying liquid volume restricts the feasibility of fixed mechanical surface aerators. The most common aeration system in SBRs are diffused bubblers; but both the floating aerator as manufactured by Aqua SBR and diffused bubble aeration systems will benefit from submerged mixers used to ensure proper agitation of the reactor contents under anoxic conditions.

As previously mentioned, the decanter is the primary piece of equipment that distinguishes different SBR manufacturers. Types of decanters include floating and fixed. Floating decanters offer the advantage of maintaining the inlet orifice slightly below the water surface to minimize the removal of solids in the effluent removed during the DRAW step. Floating decanters also offer the operating flexibility to vary fill-and-draw volumes. Fixed decanters are built into the side of the basin and can be used if the settlw step is extended. Extending the settle step minimizes the chance that solids in the wastewater will float over the fixed decanter. In some cases, fixed decanters are less expensive and can be designed to allow the operator to lower or raise the level of the decanter. Fixed decanters do not offer the operating flexibility of the floating decanters.

6.4. Health and Safety

Safety should be the primary concern in every design and system operation. A properly designed and operated system will minimize potential health and safety concerns. Manuals such as the Manual of Practice (MOP) No. 8, Design of Municipal Wastewater Treatment Plants (24), and MOP No. 11, Operation of Municipal Wastewater Treatment Plants (25) should be consulted to minimize these risks. Other appropriate industrial wastewater treatment manuals, federal regulations, and state regulations should also be consulted for the design and operation of wastewater treatment systems.

7. PROCESS PERFORMANCE

The performance of SBRs is typically comparable to conventional activated sludge systems and depends on system design and site specific criteria. Depending on their mode of operation, SBRs can achieve good BOD and nutrient removal. For SBRs, the BOD removal efficiency is generally 85% to 95% and nitrogen removal can be considerably higher than in conventional activated sludge systems (26, 27, 29–32). Performance results from full-scale facilities are provided in Table 5.6.

SBR manufacturers will typically provide a process guarantee to produce an effluent of less than (1, 3):

1. 10 mg/L BOD
2. 10 mg/L TSS
3. 5 to 8 mg/L TN
4. 1 to 2 mg/L TP

One of the primary features of SBR technology is the flexibility to exercise control as a function of time rather than space (as in conventional flow-through systems). Several key aspects include (1, 3):

1. The SBR system can tolerate shock loads and peak flows because of the equalizing basin characteristics of the fill phase.
2. Periodic effluent discharge may permit retention of reactor contents until desired clarity or treatment quality is achieved.
3. A fraction of the total volume may be used during low flow periods, resulting in lower aeration requirements. If aerators or blowers have turn-down capability, O&M costs may be reduced.
4. No RAS or internal recycles are required; however, some systems (e.g., CASS) include recycle to an antecedent basin or selector chamber.
5. With intermittently fed SBRs, clarification occurs under total quiescence, thereby eliminating short-circuiting. Consequently, small flocs will settle in an SBR that would be washed out in a continuous-flow regime.
6. Filamentous growth can be controlled by operational strategies along with adjustments during the fill phase.

Readers interested in the performance of SBR systems in industrial wastewater treatment are referred to (33–35).

Table 5.6
Summary of SBR plant operating data

Plant	Flow (m^3/day, MGD)	Influent BOD_5, (mg/L)	Influent TKN (Total N) (mg/L)	Effluent TKN (Total N) (mg/L)	Influent NH_4^+-N (mg/L)	Effluent NH_4^+-N (mg/L)	Effluent NO_x-N (mg/L)	Effluent Total N (mg/L)	% N Removal
Nonproprietary Culver, IN	N/A	170	N/A	N/A	20.0	1.0	N/A	1.0[a]	88
Cass Deep River, CT	189 (0.05)	100	54.5	3.6	40.4	1.3	1.0	4.6	92
Cass Dundee, MI	N/A	123	28.9	2.2	16.9	0.5	4.9	2.7	75
Nonproprietary Grundy Center, IA	1,249 (0.33)	210	N/A	N/A	17.3	0.8	2.8	3.6[a]	90
Aqua SBR Grundy Center, IA	3,028 (0.8)	140	28.0	4.4	19.0	1.6	0.5	4.9	83
Aqua SBR Rock Falls, IN	530 (0.14)	109	39.8	1.8	35.9	0.6	1.0	2.8	93
Aqua SBR Oak Hill, MI	416 (0.11)	220	N/A	N/A	25.0	0.6	3.5	4.1[a]	84
Jet Tech Oak Pt., MI	227 (0.06)	142	N/A	N/A	19.0	0.6	2.8	3.4[a]	82
Jet Tech Cow Creek, OK	9,841 (2.6)	119	24.0	2.7	17.0	1.8	1.9	4.6	81
Jet Tech Del City, OK	13,248 (3.5)	115	28.3	5.4	17.6	0.9	3.5	5.4	81
ICEAS Buckingham, PA	492 (0.13)	349	N/A	N/A	29.2	0.6	0.9	1.5[a]	95
ICEAS Burkeville, VA	530 (0.14)	296	35.7	3.6	19.3	0.3	1.0	4.6	87
ICEAS Shiga Kogen	757 (0.2)	484	36.9	5.4	N/A	N/A	N/A	5.4	85

Source: US EPA.
N/A—Data not available
[a] Based on effluent NH_4^+-N + N_{ox}-N.

8. OPERATION AND MAINTENANCE

The SBR typically eliminates the need for separate primary and secondary clarifiers in most municipal systems, which reduces operations and maintenance requirements. In addition, RAS pumps are not required. In conventional biological nutrient removal systems, anoxic basins, anoxic zone mixers, toxic basins, toxic basin aeration equipment, and internal MLSS nitrate-nitrogen recirculation pumps may be necessary. With the SBR, this can be accomplished in one reactor using aeration/mixing equipment, which will minimize operation and maintenance requirements otherwise be needed for clarifiers and pumps.

Since the heart of the SBR system is the controls, automatic valves, and automatic switches, these systems may require more maintenance than a conventional activated sludge system. An increased level of sophistication usually equates to more items that can fail or require maintenance. The level of sophistication may be very advanced in larger SBR wastewater treatment plants requiring a higher level of maintenance on the automatic valves and switches (1, 3). The recent advances and cost reductions of microprocessors have been some of the causes of the revival of interest in SBR technology, permitting automated control of the timing and sequence of process phases and operation. The use of timers and DO monitors can be used to reduce costs attributable to over aeration, thereby reducing the lag period of DO depletion and allowing the maximum time for denitrification to occur.

Significant operating flexibility is associated with SBR systems. An SBR can be set up to simulate any conventional activated sludge process, including BNR systems. For example, holding times in the aerated react mode of an SBR can be varied to achieve simulation of a contact stabilization system with a typical hydraulic retention time (HRT) of 3.5 to 7 hours or, on the other end of the spectrum, an extended aeration treatment system with a typical HRT of 18 to 36 hours. For a BNR plant, the aerated react mode (oxic conditions) and the mixed react modes (anoxic conditions) can be alternated to achieve nitrification and denitrification. The mixed fill mode and mixed react mode can be used to achieve denitrification using anoxic conditions. In addition, these modes can ultimately be used to achieve an anaerobic condition where phosphorus removal can occur. Conventional activated sludge systems typically require additional tank volume to achieve such flexibility. SBRs operate in time rather than in space and the number of cycles per day can be varied to control desired effluent limits, offering additional flexibility with an SBR.

9. COST

This section includes some general guidelines as well as some general cost estimates for planning purposes. It should be remembered that capital and construction cost estimates are site-specific.

Budget level cost estimates presented in Table 5.7 are based on projects that occurred from 1995 to 1998 (1). Budget level costs include such as the blowers, diffusers, electrically operated valves, mixers, biosolids pumps, decanters, and the control panel. All costs in this chapter have been updated to year 2008 costs, using the Cost Index for Utilities shown in Appendix (36). The 1998 costs were multiplied by a factor $552.16/459.40 = 1.20$ i.e., costs were increased by 20% to obtain their values in terms of 2008 USD.

Table 5.7
SBR equipment costs based on different existing facilities

Design flowrate (MGD)	Budget level equipment costs (USD)
0.012	113,000
0.015	165,000
1.0	408,000
1.4	488,000
1.46	488,000
2.0	680,000
4.25	1,408,000

Source: US EPA.
Costs are adjusted to Current 2008 USD.

Table 5.8
Equipment costs based on flowrates

Design flowrate (MGD)	Budget level equipment costs (USD)
1	182,000–422,000
5	552,000–878,000
10	1,310,000–1,649,000
15	2,648,000
20	2,528,000–3,611,000

Source: US EPA.
Costs are adjusted to Current 2008 USD.

In Table 5.8, a range of equipment costs for different design flowrates is provided (1).

Again the equipment cost items provided do not include the cost for the tanks, sitework, excavation/backfill, installation, contractor's overhead and profit, or legal, administrative, contingency, and engineering services. These items must be included to calculate the overall construction costs of an SBR system. Costs for other treatment processes, such as screening, equalization, filtration, disinfection, or aerobic digestion, may be included if required.

The ranges of construction costs for a complete, installed SBR wastewater treatment system are presented in Table 5.9 (1). The variances in the estimates are due to the type of biosolids handling facilities and the differences in newly constructed plants versus systems that use existing plant facilities. As such, in some cases these estimates include other processes required in an SBR wastewater treatment plant.

There is typically an economy of scale associated with construction costs for wastewater treatment, meaning that larger treatment plants can usually be constructed at a lower cost per gallon than smaller systems. The use of common wall construction for larger treatment systems, which can be used for square or rectangular SBR reactors, results in this economy of scale.

Table 5.9
Installed costs per gallon treated

Design flowrate (MGD)	Budget level equipment cost (USD/gal)
0.5–1.0	2.35–6.02
1.1–1.5	2.20–3.24
1.5–2.0	2.00–3.96

Source: US EPA.
Costs are adjusted to Current 2008 USD.

Operations and Maintenance (O&M) costs associated with an SBR system may be similar to a conventional activated sludge system. Typical cost items associated with wastewater treatment systems include labor, overhead, supplies, maintenance, operating administration, utilities, chemicals, safety and training, laboratory testing, and solids handling. Labor and maintenance requirements may be reduced in SBRs because clarifiers, clarification equipment, and RAS pumps may not be necessary. On the other hand, the maintenance requirements for the automatic valves and switches that control the sequencing may be more intensive than for a conventional activated sludge system. O and M costs are site specific and may range, in terms of 2008 USD, from USD 960 to USD 2410/MG (1).

10. PACKAGED SBR FOR ONSITE SYSTEMS

As discussed earlier, SBRs can be designed and operated to enhance removal of nitrogen, phosphorus, and ammonia, in addition to removing TSS and BOD. The intermittent flow (IF) SBR accepts influent only at specified intervals and, in general, follows the five-step sequence (Fig. 5.6). There are usually two IF units in parallel. Because this system is closed to influent flow during the treatment cycle, two units may be operated in parallel, with one unit open for intake while the other runs through the remainder of the cycles. In the continuous inflow SBR, influent flows continuously during all phases of the treatment cycle. To reduce short-circuiting, a partition is normally added to the tank to separate the turbulent aeration zone from the quiescent area (37).

The SBR system is typically found in packaged configurations for onsite and small community or cluster applications. The major components of the package include the batch tank, aerator, mixer, decanter device, process control system (including timers), pumps, piping, and appurtenances (37). Aeration may be provided by diffused air or mechanical devices. SBRs are often sized to provide mixing as well and are operated by the process control timers. Mechanical aerators have the added value of potential operation as mixers or aerators. The decanter is a critical element in the process. Several decanter configurations are available, including fixed and floating units. At least one commercial package employs a thermal processing step for the excess biosolids produced and wasted during the “idle” step. The key to the SBR process is the control system, which consists of a combination of level sensors, timers, and microprocessors. Programmable logic controllers can be configured to suit the owner’s needs. This provides a precise and versatile means of control.

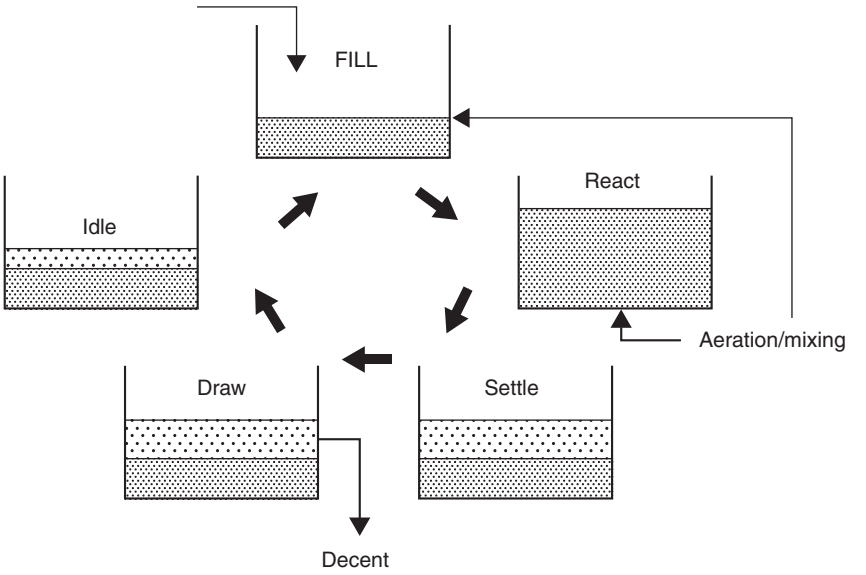

Fig. 5.6. SBR design principle for onsite systems. (*Source*: US EPA) .

10.1. Typical Applications

SBR package plants have found application as onsite systems in some states and counties where they are allowed by code. They are normally used to achieve a higher degree of treatment than a continuous-flow, suspended-growth aerobic system (CFSGAS) unit by eliminating impacts caused by influent flow fluctuations. For discharge to surface waters, they must meet effluent permit limits on BOD, TSS, and possibly ammonia, TN, and TP. Additional disinfection is required to meet effluent fecal coliform requirements. For subsurface discharge, they can be used in situations where infiltrative surface organic loadings must be reduced. There are data showing that a higher quality effluent may reduce soil absorption field area requirements. The process may be used to achieve nitrification as well as nitrogen and phosphorus removal before surface and subsurface discharge (37).

10.2. Design Assumptions

Typical IF system design information is provided in Table 5.10 (37). With CF-type (continuous flow) SBRs, a typical cycle time is 3 to 4 hours, with 50% of that cycle devoted to aeration (step 2), 25% to settling (step 3), and 25% to decant (step 4). With both types, downstream or subsequent unit processes (e.g., disinfection) must be designed for greater capacity (because the effluent flow is several times the influent flow during the decant period) or an equalization tank must be used to permit a consistent flow to those processes.

Onsite package units should be constructed of non corrosive materials, such as coated concrete, plastic, fiberglass, or coated steel. Some units are installed aboveground on a concrete slab with proper housing to protect against local climatic concerns. The units can also

Table 5.10
Design parameters for IF-type SBR systems

Parameter	SBR systems
Pretreatment	Septic tank or equivalent
MLSS (mg/L)	2,000–6,500
F/M load (lb BOD_5/day/lb MLVSS)	0.04–0.20
Hydraulic retention time (h)	9–30
Total cycle times (h)[a]	4–12
Solids retention time (day)	20–40
Decanter overflow rate (gpm/ft^2)	<100
Biosolids wasting	As needed to maintain performance

Source: US EPA.
[a]Cycle times should be tuned to effluent quality requirements, wastewater flow, and other site constraints.

be buried underground as long as easy access is provided to all mechanical parts, electrical control systems, and water surfaces. All electric components should meet NEC code and should be waterproofed and/or sheltered from the elements. If airlift pumps are used, large-diameter pipes should be provided to avoid clogging. Blowers, pumps, and other mechanical devices should be designed for continuous heavy-duty use. Easy access to all moving parts must be provided for routine maintenance. An effective alarm system should be installed to alert home owners or management entities of malfunctions (38).

10.3. Performance

With appropriate design and operation, SBR plants have been reported to produce high quality BOD and TSS effluents. Typical ranges of $CBOD_5$ (carbonaceous 5-day BOD) are from 5 to 15 mg/L. TSS ranges from 10 to 30 mg/L in well-operated systems. Fecal coliform (FC) removal of 1 to 2 logs can be expected. Normally, nitrification can be attained most of the time unless cold temperatures persist. The SBR systems produce a more reliable effluent quality than CFSGAS owing to the random nature of the wastewater generated from an individual home. The CF/SBR is also capable of meeting secondary effluent standards (30 mg/L of $CBOD_5$ and TSS), but more subject to upset by randomly generated wastewaters than the IF/SBR (39) if short-circuiting cannot be minimized.

10.4. Management Needs

Long-term management (including operation and maintenance) of SBRs through home-owner service contracts or local management programs is an important component of the operation and maintenance program. Homeowners do not typically possess the skills needed or the desire to learn to perform proper operation and maintenance. In addition, home-owner neglect, ignorance, or interference (e.g., disabling alarm systems) has contributed to operational malfunctions. No wasting of biomass should be practiced until a satisfactory concentration has developed. Intensive surveillance by qualified personnel is desirable during the first months of startup.

Most operating parameters in SBR package systems can be controlled by the operator. Time clock controls may be used to regulate cycle times for each cycle, adjusted for and depending on observed performance. Alarm systems that warn of aerator system failure and/or pump failure are essential.

Inspections are recommended three to four times per year; septage pumping (biosolids wasting) is dependent upon inspection results. Operation and maintenance requires semi-skilled personnel. Based on field experience, 5 to 12 person-hour per year, plus analytical services, are required. The process produces 0.6 to 0.9 lb TSS/lb BOD_5 removed and requires between 3.0 and 10 kWh/day for operation (37). Operating personnel prefer these systems to CFSGAS for their simplicity of O/M tasks. The key operational components are the programmer and the decanter, and these must be maintained in proper working order.

10.5. Risk Management Issues

With proper management, a package SBR system is reliable and should pose no unacceptable risks to the homeowner or the environment (37). If neglected, however, the process can result in environmental damage through production of poor-quality effluent that may pose public health risks and can result in the premature failure of subsurface systems. Odor and noise may also create some level of nuisance. SBRs are less susceptible to flow and quality loading changes than other aerobic biological systems, but they are still not suitable for seasonal applications. They are similarly susceptible to extreme cold and should be buried and/or insulated in areas subjected to these extremes. Local authorities can provide guidance on climatic effects on equipment and how to prevent them. The controller should be located in a heated environment. Long power outages can result in odors and effluent degradation, as is the case with other aerobic biological systems.

10.6. Costs

For residential applications, typical system equipment costs, in term of 2008 USD, are USD 8,560 to USD 10,700. Installation costs vary depending on site conditions; installation costs between USD 1,820 and USD 3,640 are typical for uncomplicated sites with good access (37). It should be noted that additional system components (e.g., subsurface infiltration system) will result in additional costs.

Annual operation and maintenance costs include electricity use (<USD 364/year), sludge removal (>USD 118/year), and equipment servicing. Some companies are providing annual service contracts for these units for USD 300 to USD 482 (37). Actual costs will vary depending on the location of the unit and local conditions.

REFERENCES

1. US EPA, *Sequencing Batch Reactors—Wastewater Technology Fact Sheet*, EPA 832-F-99-073, US Environmental Protection Agency, Office of Water, Washington, DC (1999).
2. E. Ardern and T. W. Lockett, Experiments on the oxidation of sewage without the aid of filters, *Journal of Society of Chemical Industry*, 33, 10, 523 (1914).

3. US EPA, Sequencing batch reactor, in *Process Design Manual for Nitrogen Removal*, EPA/625/R-93/010, US Environmental Protection Agency, Cincinnati, OH (1993).
4. Austgen Biojet Waste Water Systems, Inc., *Design Manual*, Version 1.1, No. 44, May (1991).
5. K. J. Deeny and J. A. Heidman, Implementation of sequencing batch reactor technology in the United States, *64th Annual Meeting of the Water Pollution Control Federation*, Toronto, Canada (1991).
6. R. L. Irvine, L. H. Ketchum, R. Breyfogle, and E. F. Barth, Municipal application of sequencing batch treatment, *Journal of the Water Pollution Control Federation*, 55, 484 (1983).
7. K. J. Deeney, J. A. Heidman, W. W. Schuk, K. S. Young, and A. J. Condren, Implementation of sequencing batch reactor technologies in the United States, *64th Annual Conference of the Water Pollution Control Federation*, Toronto, Ontario, October (1991).
8. US EPA, *Sequencing Batch Reactors for Nitrification and Nutrient Removal*, Report prepared by HydroQual, Inc. Washington, DC (1991).
9. ABL Environmental Consultants, *SBR Process*, http://www.ablenvironmental.com/prod_sbr_stages.htm, January (2005).
10. ECO, *Biological Nutrient Removal, SBR Sequencing Batch Reactor Nitrification Denitrification*, Located at: http://www.sequencertech.com/biotechnology/sbr/nutrient_removal.html, January (2005)
11. K. A. Mikkelson, *Aqua SBR Design Manual*, Aqua-Aerobic Systems, Copyright (1995).
12. US Filter, *OMNIFLO Sequencing Batch Reactor (SBR) System*, Located at http://www.usfilter.com, January (2005).
13. Fluidyne, *SBR Sequencing Batch Reactor*, http://www.fluidynecorp.com/sbr_seqbatch.html (2004)
14. Aqua Aerobics, Babcock King-Wilkinson, L. P., *Manufacturers Information*, Fluidyne, and Jet Tech Systems (1998).
15. M. C. Goronszy, *CASS, Cyclic Activated Sludge System: Superior Batch Reactor Technology*, Transenviro, Inc. Aliso Viejo, CA (1990).
16. W. M. Shubert, *SCR-Sequencing Batch Reactor*, Design manual for Aqua-Aerobic Systems, Inc (1986).
17. K. J. Deeny, J. A. Heidman, W. W. Schuk, K. S. Young, and A. J. Condren, Implementation of sequencing batch reactor technologies in the United States, *64th Annual WPCF Conference*, Toronto, Ontario, (1991).
18. US EPA, *Sequencing Batch Reactors*, Summary report for the Center for Environmental Research Information, EPN625/8-86/011, US Environmental Protection Agency, Cincinnati, OH (1986).
19. M. C. Goronszy, Nitrogen removal and sludge bulking control in cyclically operated activated sludge systems, *Proceedings of the 61st Meeting of Ohio Water Pollution Control Association*, Akron, OH (1987).
20. G. T. Daigger, M. H. Jr. Robbins, and B. R. Marshall, The design of a selector to control low F: M filamentous bulking, *Journal of the Water Pollution Control Federation*, 57, 3 (1985).
21. P. A. Wilderer, R. L. Irvine, and M. C. Goronszy, *Sequencing Batch Reactor Technology*, IWA Publishing, London, March (2001).
22. E. M. Toby, III, Sequencing Batch Reactor (SBR) Operations and Trouble Shooting, Training Course, Treeco Center, University of Florida, Can be found at http://www.treeo.ufl.edu/water/courses.asp?course=1472 (2005).
23. Parsons Engineering Science, Inc., *Basis of Design Report—Urgent Extensions to Maray Sewer Treatment Works*, Abu Dhabi, UAE (1992).

24. WEF, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice (MOP) No. 8, Water Environment Federation, Alexandria, VA (1998).
25. WEF, *Operation of Municipal Wastewater Treatment Plants*, Manual of Practice (MOP) No. 11, Water Environment Federation, Alexandria, VA (2007).
26. S. S. Choi and Y. J. Yoo, *Removal of phosphate in a sequencing Batch Reactor by Staphylococcus auricularis, Biotechnology Letters*, 22, 19, 1549–1552, October (2000).
27. S. R. Linne and S. C. Chiesa, Operational variables affecting performance of the selector—complete mix activated sludge process, *Journal of the Water Pollution Control Federation*, 59, 7 (1987).
28. N. H. Callado and E. Foresti, Removal of organic carbon, nitrogen and phosphorus in sequential batch reactors integrating the aerobic/anaerobic processes, *Water Science & Technology*, 44, 4, 263–270 (2001).
29. Pennsylvania WRRC, *Nutrient Removal of a Sequencing Batch Reactor (SBR) Treating Wastewater with Potential for Water Reclamation*, Pennsylvania Water Resources Research Center (WRRC), FY 2004–2005 Research Projects http://www.pawatercenter.psu.edu/research_projects/04_05_sbr_li.htm (2004).
30. J. A. D. Rodrigues, A. G. Pinto, S. M. Ratuszznei, M. Zaiat, and R. Gedraite, Enhancing of the performance of an anaerobic sequencing batch reactor treating low-strength wastewater, through implementation of a variable Stirring rate program, *Brazilian Journal of Chemical Engineering*, 21, 3, 423–434 (2004).
31. L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas, *Biological Treatment Processes*, Humana Press, Totowa, NJ (2009).
32. L. K. Wang and Y. Li, Sequencing batch reactors, In *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas, (eds.), Humana Press, Totowa, NJ (2009).
33. D. I. Masse, Treatment of slaughterhouse wastewater in anaerobic sequencing batch Reactor, *Canadian Agricultural Engineering*, 42, 3, 131–137 (2000)
34. B. M. Kumar and S. Chaudhari, Evaluation of sequencing batch reactor (SBR) and sequencing batch biofilm reactor (SBBR) for biological nutrient removal from simulated wastewater containing glucose as carbon source, *Water Science and Technology*, 48, 3, 73–79 (2003).
35. D. P. Whichard, Nitrogen Removal from Dairy Manure Wastewater Using Sequencing Batch Reactors, M.S. thesis, Virginia Polytechnic Institute and State University, Blacksburg, VA, July (2001).
36. US ACE, *Civil Works Construction Cost Index System Manual*, 110-2-1304, US Army Corps of Engineers, Washington, DC, pp. 44, http://www.nww.usace.army.mil/cost, (2008).
37. US EPA, *Sequencing Batch Reactor Systems*, Onsite Wastewater Treatment Systems Technology Fact Sheet 3, EPA 625/R-00/008, US Environmental Protection Agency, Office of Research and Development, National Risk Management Research Laboratory (2000).
38. A. S. Eikum and T. Bennett, New Norwegian technology for treatment of small flows, *Proceedings of Seventh Northwest Onsite Wastewater Treatment Short Course*, R. W. Seabloom (ed.), University of Washington, Seattle (1992).
39. Ayres Associates, *Florida Keys Onsite Wastewater Nutrient Reduction Systems Demonstration Project*, HRS Contract No. LP988, Florida Department of Health, Gainesville, FL (1998).

APPENDIX

United States Yearly Average Cost Index for Utilities US Army Corps of Engineers (36)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

6
Simultaneous Nitrification and Denitrification (SymBio® Process)

Hiren K. Trivedi

CONTENTS

Abstract Simultaneous nitrification and denitrification (SND) is an attractive option for design engineers and scientists as it may offer significant advantages compared to conventional processes with separate nitrification and denitrification reactors. SND eliminates the need for a separate denitrification tank and mixed liquor recycle. The SymBio® process uses measurement of the intracellular pool of reduced nicotinamide adenine dinucleotide (NADH) for assessing the real-time biological activity in activated sludge systems. This information is used to control the air supply in the aeration tank to maintain DO at the desired low level, which ensures that both anoxic and aerobic zones are developed in sludge flocs. This allows nitrification and denitrification to occur simultaneously in the same reactor. The SymBio® process control concept has also been combined with MBR technology in the US wastewater treatment industry.

Key Words Activated sludge • BNR • denitrification • fluorescence • MBR • NADH • nitrification • SND.

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_6 © Humana Press, New York, NY 2009

1. INTRODUCTION

Clean water is a priority in our society. Biological nutrient removal (BNR) is becoming increasingly common in both domestic and industrial wastewater treatment. Innovative BNR processes have been developed and applied to full-scale wastewater treatment plants in the last two decades. Nitrogen and phosphorous are the major nutrients of concern as they promote eutrophication of natural water systems and stimulate growth of algae. Although the removal of phosphorous can be achieved both chemically and biologically, nitrogen removal is usually performed biologically.

The biological transformations of nitrogen are comprised of six major processes (1):

1. *Assimilation* of inorganic forms (primarily ammonia and nitrate) by plants and microorganisms to form organic nitrogen, e.g., amino acids, proteins, purines, pyrimidines, and nucleic acids.
2. *Heterotrophic conversion* of organic nitrogen from one organism (food or prey) to another organism (consumer or predator).
3. *Ammonification*, the decomposition of organic nitrogen to ammonia.
4. *Nitrification*, the oxidation of ammonia to nitrite and nitrate.
5. *Denitrification*, the bacterial reduction of nitrate to molecular nitrogen (N_2).
6. *Nitrogen fixation*, the reduction of nitrogen gas to ammonia and organic nitrogen by various microorganisms.

As far as wastewater treatment is concerned, total nitrogen is comprised of organic nitrogen, ammonia, nitrite, and nitrate. Removal of nitrogen from wastewater is desirable as some forms of nitrogen can cause problems if they are discharged to the environment untreated. For example, ammonia is toxic to fish and it can deplete the oxygen resources. Nitrate may cause potential adverse health effects, including methemoglobinemia (a reduction in the oxygen-carrying capacity of the blood) in infants and nitrite can cause cancer in animals through formation of N-nitroso compounds. This chapter discusses a commercially available process (the SymBio® process) designed for biological nitrogen removal using simultaneous nitrification and denitrification in wastewater treatment plants.

2. BIOLOGICAL NITROGEN REMOVAL

In BNR plants designed for nitrogen removal, the bacterial mass is alternatively exposed to conditions of oxygen abundance and oxygen shortage. The differing oxygen concentrations promote the biological activity of one or more groups of bacteria and distinguish different phases in the wastewater treatment process. These phases can be spatially separated, with the sludge circulating between tanks or zones maintained at differing oxygen concentrations. Alternatively, the phases can also be separated in time, so that the sludge remains in a single tank, e.g., in a sequential batch reactor, whereas the oxygen concentration is varied in a controlled manner using a timer.

In both types of installations, the wastewater is brought into contact with the sludge so that the pollutants are reduced to harmless substances. Ammonia is oxidized to nitrite and then to nitrate by nitrifying bacteria in presence of oxygen during the nitrification phase. The nitrate is subsequently reduced to molecular nitrogen (N_2) by the denitrifying bacteria. Organic matter is oxidized in presence of oxygen or nitrate, acting as electron acceptors. Switching between

different phases can be viewed as changes between various metabolic conversion paths, which cumulatively result in production of new bacterial biomass, CO_2, H_2O, and N_2 (2).

2.1. Nitrification

The two-step oxidation of ammonia to nitrate is performed by autotrophic nitrifying bacteria, *Nitrosomonas* and *Nitrobacter*, as shown below (3):

Nitrosomonas

$$2NH_4^+ + 3O_2 \rightarrow 2NO_2^- + 4H^+ + 2H_2O \quad (1)$$

Nitrobacter

$$2NO_2^- + O_2 \rightarrow 2NO_3^- \quad (2)$$

Total Reaction

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O \quad (3)$$

Based on this reaction the oxygen requirement for complete oxidation of ammonia to nitrate is approximately 4.57 g of O_2 per gram of ammonia-N oxidized. Further, approximately 7.14 g of alkalinity (as $CaCO_3$) are consumed per gram of ammonia-N oxidized.

It should be noted that the *US EPA Nitrogen Control Manual* (4) provides the following equation for nitrification, which accounts for both synthesis and oxidation:

$$NH_4^+ + 1.83O_2 + 1.98HCO_3^- \rightarrow 0.21C_5H_7O_2N + 0.98NO_3^- + 1.041H_2O + 1.88H_2CO_3 \quad (4)$$

Based on this equation, the oxygen requirement for ammonia oxidation is lower, approximately 4.2 g/g ammonia-N oxidized.

Although nitrification in wastewater treatment is primarily attributed to *Nitrosomonas* and *Nitrobacter*, Wagner et al. (5) recently showed, using oligonucleotide probes, that *Nitrococcus* was dominant species for nitrite oxidation instead of *Nitrobacter*.

2.2. Denitrification

For certain bacteria nitrate and nitrite are both strong oxidizing agents and potential sources of nitrogen. Consequently, different groups of bacteria exploit them in different ways. In assimilative nitrate reduction, nitrate is reduced to ammonia for use as nitrogen source for growth. In dissimilative nitrate reduction (e.g., denitrification), nitrate is used as an alternative electron acceptor in energy generation. Assimilative nitrate reduction occurs in all plants and most fungi, as well as in many bacteria, whereas dissimilative nitrate reduction or denitrification is restricted only to bacteria. Denitrification involves reduction of nitrate to nitrite to nitric oxide to nitrous oxide to nitrogen gas:

$$NO_3^- \rightarrow NO_2^- \rightarrow NO \rightarrow N_2O \rightarrow N_2 \quad (5)$$

Denitrification is performed by several heterotrophic bacteria including *Achromobacter, Acinetobacter, Agrobacterium, Alcaligenes, Arthobacter, Bacillus, Chromobacterium,*

Corynebacterium, Flavobacterium, Hypomicrobium, Moraxella, Nesseria, Paracoccus, Propionibacterium, Pseudomonas, Rhizobium, Rhodopseudomonas, Spirillum, and *Vibrio* (6).

As nitrate is used as a terminal electron acceptor instead of oxygen during denitrification, the aeration requirement for biochemical oxygen demand (BOD) is reduced. The oxygen equivalent of nitrate nitrogen in oxidation-reduction reactions is 2.86 g/g of nitrate-N. Further, 3.57 g of alkalinity (as $CaCO_3$) is recovered per gram of nitrate-N denitrified (3).

As mentioned before, ammonia oxidation to nitrate (nitrification) requires a source of oxygen (aerobic environment) whereas optimum denitrification requires absence of oxygen, along with presence of organic carbon, to require the use of nitrate as an electron acceptor for energy generation (anoxic environment). Subsequently, conventional BNR systems designed for nitrogen removal usually include two separate reaction zones. These are created in separate tanks or in separate sections (zones) within a tank. They can also be created in separate cycles in a sequential batch mode. In a configuration that involves separate tanks (or zones), the wastewater initially enters a denitrification (pre-anoxic) zone to which nitrified "mixed liquor" is recycled from a subsequent nitrification compartment. BNR processes like MLE, UCT, Bardenpho, and A^2O use configurations that involve pre-anoxic steps.

The denitrification environment can also be created after the nitrification step, as a post-anoxic step. This was a common configuration in 1970s. However, since denitrification is primarily a heterotrophic reaction, availability of organic carbon is a major requirement for it to proceed successfully. The importance of carbon availability and the necessary minimum BOD to nitrogen ratio required for effective denitrification are discussed later on in this chapter. In systems involving post-anoxic steps, BOD depletion during the aerobic nitrification steps creates a need for a supplemental organic carbon source (e.g., methanol) in the post-anoxic steps. Hence, configurations involving post-anoxic step, where organic carbon is externally added, are less common nowadays.

2.3. Simultaneous Nitrification and Denitrification

Recent studies have revealed that nitrification and denitrification can also occur concurrently in the same reactor. This phenomenon is called simultaneous nitrification and denitrification (SND).

SND is an attractive option for design engineers and scientists as it may offer significant advantages compared to conventional processes with separate nitrification and denitrification reactors. For example, SND eliminates the need for a separate denitrification tank and mixed liquor recycle. The phenomena of SND can be explained with the following three hypotheses (7, 8):

- Anoxic/oxic zones within a sludge floc-microscopic environment

Activated sludge floc can contain both aerobic and anoxic zones. Depending on the dissolved oxygen (DO) concentration and concentration of BOD and ammonia, oxygen may be depleted towards the center of the floc. This means that oxygen cannot diffuse through the entire floc depth and results in oxygen gradient across the floc. This will allow the nitrate generated in the outer, aerobic zone to diffuse into this inner, anoxic zone along with substrate so that denitrification occurs simultaneously.

- Anoxic/oxic zones within a bioreactor-macroscopic environment

Regions of low DO or zero DO can develop within the bioreactor as a result of mixing and aeration patterns. This is particularly true for basins with surface aerators where DO depletion in regions away from these aerators is common. This allows nitrification and denitrification to occur concurrently in a single reactor.

- Presence of novel microorganisms

Certain microorganisms can contribute towards nitrogen removal in a single reactor. For example, Robertson et al. (9) indicated that *Thiosphaera pantotropha*, a heterotrophic organism, could simultaneously nitrify and denitrify under aerobic conditions. Davies et al. (10) provided evidence for aerobic denitrification for *Pseudomonas aeruginosa* and *Paracoccus denitrificans*. Certain autotrophic organisms are also known to have denitrification capabilities.

The objective here is to discuss the SymBio® process that can maintain conditions for SND by controlling the development of the anoxic and the aerobic zones within sludge flocs at microscopic level. This chapter describes how measurement of the intracellular pool of reduced nicotinamide adenine dinucleotide (NADH) represents an effective means of assessing the real-time biological activity in the SymBio® process. With this information, it is possible to decide whether the biological process is in a state of balance or imbalance. In the SymBio® process, this information is used to control the air supply in the aeration tank to maintain DO at the desired low level, which ensures that both anoxic and aerobic zones are developed in sludge flocs. This allows simultaneous nitrification and denitrification in the same reactor. Before discussing the SymBio® process in detail, it is necessary to explain the role of NADH in bacterial metabolism, which is described below.

3. NADH IN CELL METABOLISM

Many fluorophores, both intracellular and extracellular, are present in biological processes including the activated sludge systems. The concentration variations of fluorophores are often closely related the cell activities and, therefore, can be used as indicators of important process parameters such as cell concentration, metabolic stage, growth, death, etc. The best-studied biological fluorophores are nicotinamide adenine dinucleotides. All living cells contain coenzymes nicotinamide adenine dinucleotide (NAD) and nicotinamide adenine dinucleotide phosphate (NADP), which serve as major electron/hydrogen carriers in oxidation-reduction reactions of metabolism. Accompanying substrate catabolism, the oxidized form of coenzyme, NAD^+, which contains a reactive pyridine ring, is reduced in position 4 by a hydride ion to form NADH. NADH is directly involved in ATP generation via oxidative phosphorylation in respiration. Through these processes, the reduced form of coenzyme NADH is oxidized back to NAD^+ as shown in Fig. 6.1. Thus, NADH assumes a central position in the internal microbial energy transport. The NADH/NAD^+ cycle is described in detail below:

$$NAD^+ + H^+ + 2e^- \rightarrow NADH(\text{Substrate Catabolism}) \quad (6)$$

$$NADH \rightarrow NAD^+ + H^+ + 2e^-(\text{Respiration}) \quad (7)$$

NADH is produced in large quantities during the oxidation of carbohydrates like glucose. Glycolysis is a method of decomposition commonly used by bacteria for breaking down

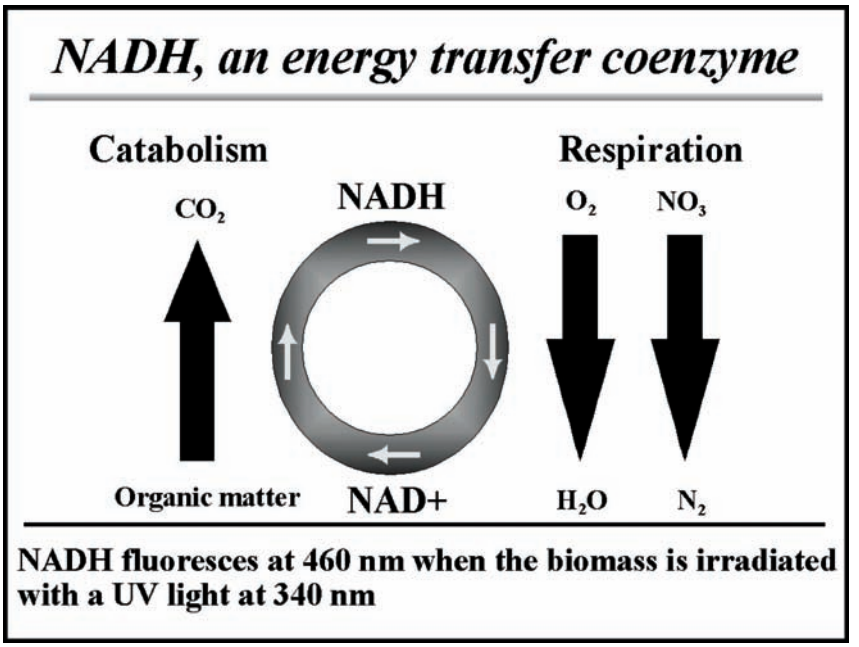

Fig. 6.1. Cyclic nature of NADH in bacterial metabolism.

carbohydrates. Glycolysis yields 2 moles of NADH for each mole of glucose converted. Two moles of the energy-rich, phosphorylated compound ATP are formed at the same time. The reactions that form ATP in this manner are known as substrate-level phosphorylation. ATP can also be formed by electron-transport phosphorylation during the oxidation of NADH to NAD^+. This oxidation of NADH is carried out by a large number of enzymes that are embedded in the cell membranes of the bacteria. These enzymes constitute a respiratory chain where oxygen is usually used as a terminal electron acceptor. Part of the NADH formed during glycolysis and the tricarboxylic acid cycle is oxidized to NAD^+, catalyzed by the enzymes of the respiratory chain and combined with the production of large quantities of ATP as shown in Fig. 6.2. The ATP is subsequently used by the bacterial cells to synthesize new cell materials. Only a part of the NADH formed during substrate-level phosphorylation is oxidized in the respiratory chain. The remainder is used directly for the purpose of synthesis (11).

Certain microorganisms can also use oxidants such as nitrate (and also the sulfate, carbonate, and even other organic compounds) as terminal electron acceptors when oxygen is not available for respiration. Thus, denitrification achieved in the anoxic stage of a BNR process, operates with a reduction-oxidation cycle of NADH very similar to that of aerobic respiration, with nitrate replacing the role of oxygen (Fig. 6.1).

Cycles of reduction-oxidation for coenzyme NADH also exist in fermentation (anaerobic condition), which is not shown in Fig. 6.1. In fermentation, no externally supplied electron acceptor is required. The generation of NAD^+ from NADH is coupled with subsequent

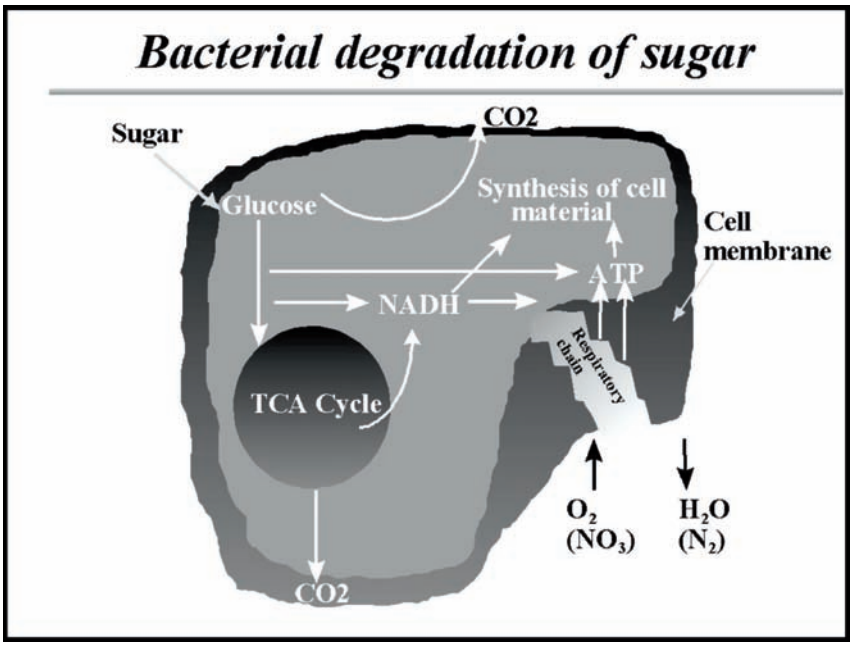

Fig. 6.2. Generalized schematic representation of heterotrophic metabolism. (*Source*: (11)).

reduction of an oxidized organic compound (e.g., acetyl-CoA or pyruvate in the case of *Escherichia coli*) generated from catabolism of the initially fermentable substrate (12).

The concentration of NADH in a living bacterial cell is determined by the balance between the rates of reduction (generation) and oxidation (consumption) reactions. The oxidizing power of the organic compounds in the oxidation of NADH in fermentation is much weaker than those of nitrate and oxygen. For example, the reduction potential for the oxidation-reduction pair of pyruvate/lactate is -0.19 V, whereas those for NO_3^-/N_2 and $^1/_2O_2/H_2O$ are $+0.74$ V and $+0.82$ V respectively (13). Consequently, the rate of NADH oxidation is much slower with anaerobic metabolism than with denitrification and aerobic respiration. The intracellular level of NADH at anaerobic stage is therefore higher than those at anoxic or oxic stages. Further, as reduction potential for the oxidation-reduction pair NO_3^-/N_2 is lower than that for $^1/_2O_2/H_2O$, the NADH level is higher under anoxic condition than under aerobic conditions. This is simply because oxygen, with higher oxidizing power, oxidizes the intracellular NADH to a lower level than nitrate does. Also, the microorganism population in a wastewater treatment plant is a combination of many microbial species. Since not all the species are capable of utilizing nitrate as the terminal electron acceptor, a portion of the population does not respond to a shift from anaerobic to anoxic conditions (12). As a result, distinct difference in the level of intracellular NADH is observed under various metabolic conditions and rapid increase in NADH level is expected as biomass switches from aerobic to anoxic to anaerobic metabolism (Fig. 6.3).

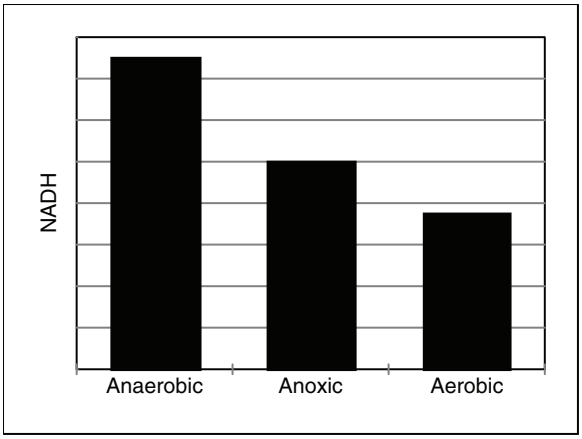

Fig. 6.3. NADH concentration under various bacterial metabolic conditions.

The NADH molecule has a molecular weight of 663 and is soluble in water. The concentration of NAD^+ in bacterial cells is of the order of 10^{-3} M. Under steady state conditions, the concentration of NADH is of the order of 10^{-6} M, i.e., the ratio of NAD^+ to NADH is about 1000:1. Thus, a 10% reduction in the concentration of NAD^+ will be reflected in a hundredfold increase in the concentration of NADH. Further, NADH (and not NAD^+) absorbs light at a wavelength of 340 nm and fluoresces at a wavelength of 460 nm (11). The absorption of the light is due to the fact that electrons in the NADH molecules are excited and receive a quantum of energy corresponding to the energy of the photon that was absorbed. However, the duration of this high energy state is extremely short (10^{-9} s), and light is emitted at a lower wavelength. This phenomenon is known as "fluorescence." Maximum fluorescence for NADH occurs at 460 nm, which is visible blue light. This property offers an excellent opportunity for measuring NADH concentration through measuring the level of fluorescence. This emission light at 460 nm is measured by a sensor and converted to a 4–20 mA signal (2).

NADH fluorescence monitoring differs in one significant aspect from other methods of monitoring used in wastewater treatment plants, as it monitors the conditions prevailing within the sludge flocs. For instance, oxygen and redox sensors monitor the conditions in the free water phase between the sludge flocs. From the process standpoint, the limitation of these methods is that the oxygen concentration measured in the aeration tank does not have any direct relationship to the concentration of oxygen in the sludge flocs, since the oxygen penetration of the flocs is dependent upon the consumption of oxygen within them (11).

4. THE SYMBIO® PROCESS FOR SIMULTANEOUS NITRIFICATION AND DENITRIFICATION

The objective in SymBio® process is to maintain a dual-zone phenomenon within a sludge floc where the outer region of the floc has access to the dissolved oxygen and is capable of nitrification but the inner core is oxygen depleted and is maintained under anoxic

(denitrifying) condition. This allows simultaneous nitrification and denitrification in a single floc and consequently in a single tank. As described before, NADH fluorescence is monitored within an aeration tank using a sensor and any variation in the signal intensity is used to adjust the airflow rate to maintain low dissolved oxygen below 1.0 ppm. All the cells exposed to the 340 nm ultraviolet light are monitored for NADH activity.

There are a couple of ways to use the NADH fluorescence signal to control the air supply to maintain SND with SymBio® process. They are described in detail below.

4.1. NADH Proportional Control Strategy

As it has been discussed previously, the NADH fluorescence is strongly influenced by the oxygen and nitrate concentration. Further, at a constant oxygen concentration in the water phase between the flocs, the depletion of nitrate and oxygen and as a consequence, the accumulation of NADH inside the sludge flocs is strongly dependent on the organic loading rate. To keep an optimum balance between the nitrification rate and the denitrification rate, NADH can be used to control the oxygen set point.

The bacteria in the aeration tank of a wastewater treatment plant are not evenly distributed, but flocculate in sludge flocs between which there is a free water phase. This means that the bacteria do not all have equal access to substrate and hydrogen acceptors, such as oxygen and nitrate. Although sufficient oxygen is dissolved in the water phase, the bacteria that are outermost in the sludge flocs will be well supplied with oxygen. On the other hand, bacteria that are closer to the center of the sludge flocs may have limited access to oxygen, as the oxygen concentration is determined by the total effect of the diffusion resistance and the oxygen consumption in the layer between the surfaces of the sludge flocs and the bacteria.

A simple model splits the sludge flocs into an anoxic inner core and an aerobic external shell (Fig. 6.4). If the oxygen concentration of the free water phase drops or if the oxygen

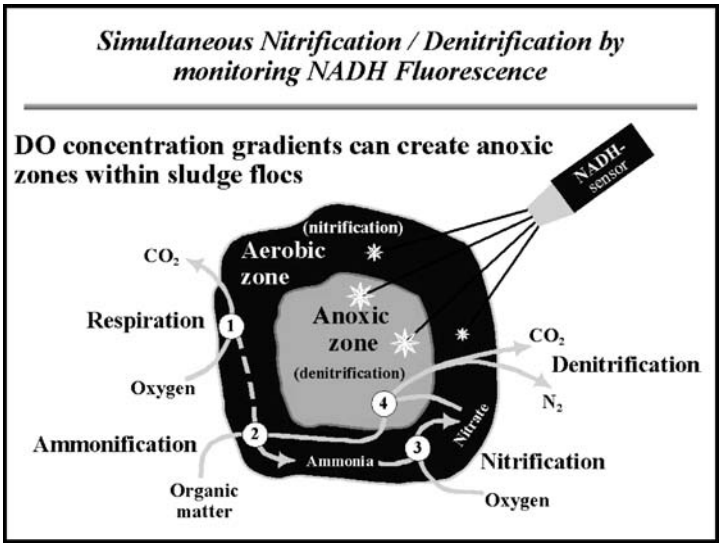

Fig. 6.4. Aerobic and anoxic regions within a sludge floc in the simultaneous nitrification and denitrification system. (*Source*: (11)).

consumption of the sludge flocs rises, for example due to an increase in the organic loading rate, the core becomes larger and the shell becomes thinner. This means that the total mass of bacteria that have a high concentration of NADH increases (Fig. 6.3). It also means that there is less filtering of the sensor's UV light as the outermost aerobic shell becomes thinner. The same applies to filtering of the fluorescence emitted by the NADH in the same layer. Thus, all of these phenomena increase the level of fluorescence detected by the sensor when the concentration of oxygen dissolved in the free water phase drops (2). Such increase in NADH fluorescence is used to automatically increase the airflow to the system.

An increase in the concentration of oxygen in the free water phase or a drop in the sludge flocs' oxygen consumption, due to a reduction in the organic loading rate, causes the oxygen to diffuse further into the flocs. This gives more bacteria the opportunity to oxidize NADH through the respiratory chain, and the quantity of NADH in the sludge flocs drops. Armed with this information, it is possible to decide whether the biological process is in a state of balance or imbalance. This knowledge can then be put to immediate use to control one or more critical process parameters such as the level of aeration, the rate of sludge return, the MLSS concentration or the end of the denitrification phase. For example, in the NADH proportional control strategy, a decrease in NADH fluorescence, as described above, is used to automatically reduce the airflow to the system.

Figure 6.5 indicates the effects of organic loading changes or the DO variations on the level of NADH and hence on its fluorescence intensity. This information is effectively used

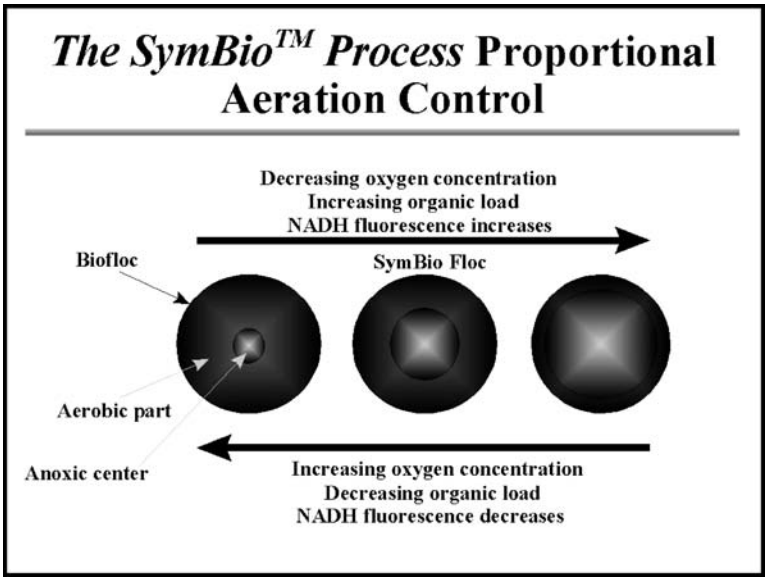

Fig. 6.5. Control of air requirements using NADH fluorescence in the simultaneous nitrification and denitrification system. (*Source*: (11)).

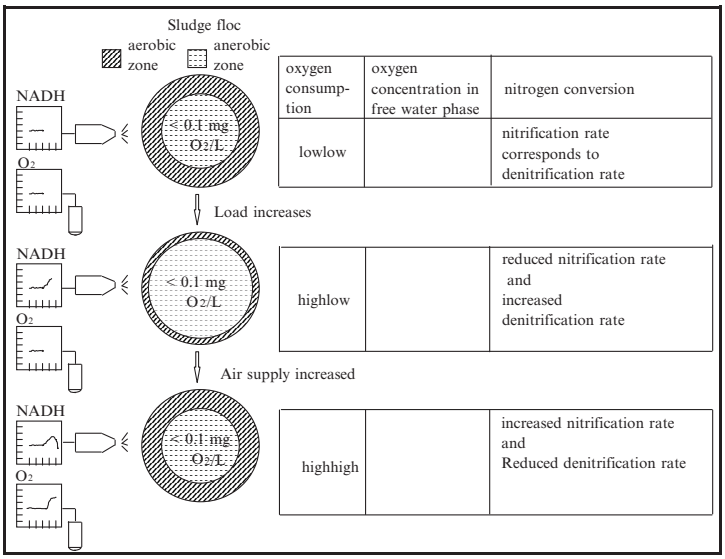

Fig. 6.6. NADH dependent proportional control strategy of the simultaneous nitrification and denitrification Process. (*Source*: Reference (11)).

in the NADH Proportional Control Strategy for air adjustments. The influence of NADH Proportional Control Strategy on the anoxic and aerobic zones of the sludge flocs upon an increment of the organic loading rate is illustrated in Fig. 6.6 (11).

As described above, NADH measurement provides an effective tool for monitoring the changes in the oxygen demand of the biomass. The NADH fluorescence measured by the sensor is converted to a 4–20 mA signal and is used as an input to a programmable logic controller (PLC). The PLC in turn proportionally controls the air supply from the blowers or surface aerators as described before. However, the concentration of NADH in the biomass is affected by changes in the mixed liquor suspended solids (MLSS) as well as the operating temperature in the reactor. For example, higher MLSS represents more biomass, which in turn means a higher NADH pool under any given metabolic condition. In contrast, an increase in temperature actually decreases the florescence intensity. So certain environmental parameters affect the NADH fluorescence intensity, which means that periodic recalibration of NADH operating range is required if only the NADH proportional control strategy is used in a SymBio® system (11).

4.2. NADH Jump Control Strategy

Where interference from the external parameters, such as described in the previous section is a factor, the NADH Jump Control Strategy is utilized. Air is cycled between a "high" DO phase (<1.0 ppm) and a "low" DO phase (<0.2 ppm). The high DO phase promotes higher nitrification and low DO phase favors denitrification. The high DO phase is of fixed length but the low DO phase is of a variable length. Any significant increase in NADH during the

low DO phase indicates a shift from relatively aerobic/anoxic conditions to purely anaerobic conditions (nitrate depletion) and this NADH increase or "jump" is used to terminate the low DO phase and initiate the next high DO phase automatically. Very low concentrations of total nitrogen can be achieved with the NADH jump control strategy when the influent BOD to nitrogen ratio is high and favorable for denitrification. As NADH levels between two consecutive phases are compared at any given time, this strategy is relatively independent of any other interference such as MLSS and temperature. Further, by controlling DO below 1.0 ppm, simultaneous nitrification and denitrification is continuously maintained.

An average NADH value is taken during a high DO (nitrification) phase. This is added to a plant specific constant (NADH constant) to establish an NADH set point for the subsequent low DO (denitrification) phase. If the actual NADH value in the denitrification phase exceeds the set point, it indicates nitrate depletion and the PLC program automatically terminates the low DO phase and initiates the next high DO phase. A smaller NADH constant results in a lower NADH set point and allows a quicker termination of the denitrification phase, which can be used if ammonia removal is the main objective. A larger constant allows complete nitrate depletion before the set point is exceeded (14).

Figures 6.7, 6.8, and 6.9 demonstrate NADH Jump Control Strategy employed in an industrial pilot plant. The loading into the aeration tank is 2000+ ppm COD and 100+ ppm TKN. The effluent nitrate ranged from 150–200 ppm before the SymBio® application. Due to chemical nature of waste, the sludge age was maintained in range of 20–25 days with water temperature at 25°C.

In Fig. 6.7, the high DO phases and the Low DO phases are of 60 min each. However, to indicate the importance of NADH monitoring, a small NADH constant (Biological Potential Unit, BPA) was chosen. Because of the smaller set point, NADH value exceeds the set point

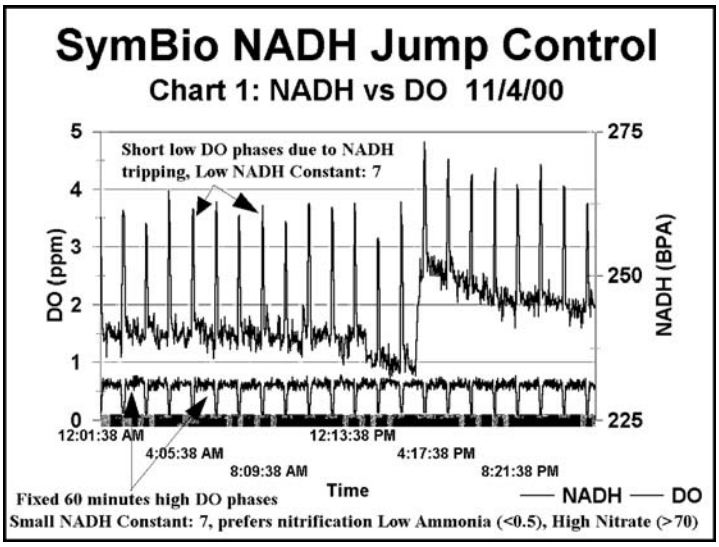

Fig. 6.7. NADH Jump Control Strategy—Chart 1. (*Source*: (14)).

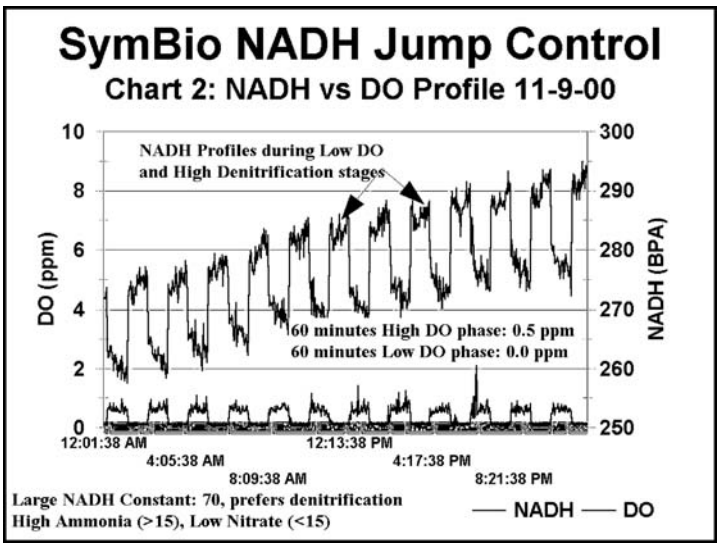

Fig. 6.8. NADH Jump Control Strategy—Chart 2. (*Source*: (14)).

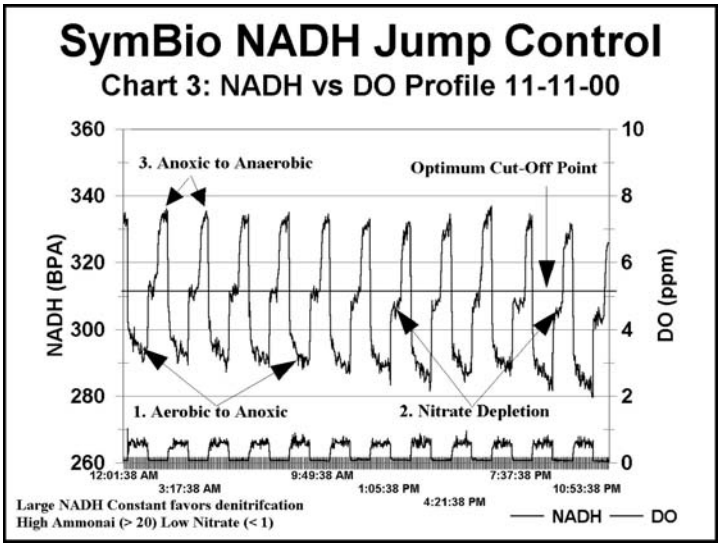

Fig. 6.9. NADH Jump Control Strategy—Chart 3. (*Source*: (14)).

quickly (<10 min), once the system is switched to the low DO phase. This terminated the low DO phase before the scheduled 60 min and hence nitrate was never completely depleted. This favored nitrification and kept ammonia below 1.0 ppm. Nitrate was reduced to around 70 ppm because of simultaneous denitrification even in the high DO phase. As Fig. 6.7 indicates, a sudden increase in the NADH level was observed around 4.00 p.m. This could have been

due to an influx of higher organic loading. Subsequently, the overall NADH concentration remained at a higher level from 4.00 p.m. to 12.00 a.m. in Fig. 6.7. However, as mentioned before, the NADH jump control strategy was not affected by this sudden increase in the absolute NADH fluorescence value as two consecutive phases of high DO and low DO were compared at any given time (14).

Figure 6.8 indicates that the NADH constant was increased to 70 BPA to favor denitrification. This allowed the low DO phase to extend to the full 60 min. Although nitrate was still not completely depleted, its effluent concentration dropped to less than 15 ppm. Every time the DO was lowered, the NADH jumped only once at the start of the low DO phase indicating that anoxic conditions prevailed and the biomass never switched to anaerobic metabolism. The overall rising trend in NADH could have been due to changes in organic loading, MLSS, or temperature. Again, as the difference in the NADH concentration between two consecutive two phases was compared, the control system was not affected (14).

As shown in Fig. 6.9, by allowing gradual nitrate depletion, a stage was reached when nitrate was eventually depleted from the system and the NADH fluorescence showed two jumps during all low DO phases. First jump indicated a switch from more aerobic to more anoxic conditions and the second jump indicated a switch from anoxic to anaerobic conditions. This was confirmed with the chemical analysis of the effluent 24-h composite samples. This demonstrated the capability of the SymBio® system using the NADH jump control strategy to switch from complete nitrification to complete denitrification conditions by adjustments of the operating parameters (14).

In wastewater treatment plants treating domestic sewage, the SymBio® system utilizing the NADH jump control strategy has been successfully applied to achieve effluent total nitrogen levels below 5.0 ppm and ammonia-N levels below 1.0 ppm. The process design considerations are discussed before the operational data from various installations is described.

4.3. Process Design

For a wastewater treatment plant operating in a conventional nitrification mode, the SymBio® process offers significant advantages such as:

- Total nitrogen removal without any concrete modifications
- Energy savings due to reduced aeration requirement
- Alkalinity recovery due to denitrification
- Reduced sludge production due to lower anoxic sludge yield

The major requirements for a simultaneous nitrification and denitrification operation are (a) control over the air supply and (b) sufficient sludge age to ensure complete nitrification and denitrification at low DO.

The energy savings compared to a nitrification system are attributed to: low DO operation; higher oxygen transfer efficiency for a given aeration system operating at low DO concentration, e.g., 0.5 ppm DO in the SymBio® mode versus 2.0 ppm DO in a conventional nitrification mode; and use of nitrate for organic carbon removal instead of oxygen. Higher oxygen transfer efficiency results from the fact that the driving force, for transferring the gaseous oxygen in

the air to the liquid water, is higher when the DO concentration in the receiving water is low. For example, at 20°C and 760 mm Hg barometric pressure, the maximum solubility of oxygen in water is 9.08 ppm (15). Hence, under these conditions, the concentration gradient for transferring oxygen into the water with 2.0 ppm DO is lower than that for the water with 0.5 ppm DO (7.08 ppm versus 8.58 ppm). This results in higher oxygen transfer efficiency for the low DO operation. Twenty-five percent to 30% energy savings have been achieved in some cases when nitrification systems were converted to SymBio® systems as described in the case studies later on in this chapter.

To achieve maximum energy savings and to maintain a balance between aerobic/anoxic sections of the sludge flocs, a good aeration control system is necessary. Typically, an NADH monitoring system coupled with variable frequency drive (VFD) on a surface aerator motor or a positive displacement blower motor is used. Inlet valve with automatic control can also be used for a centrifugal blower. One of the important requirements, in some cases, for the SymBio® process is the need for a separate mechanical mixing to avoid settling of solids in the aeration basin when operating at a low DO. In a SymBio™ system, a separate mixing device is recommended if the aeration requirement falls below 15 SCFM per 1000 ft^3 of the tank volume. Typical mixing requirement is 25 hp mixing energy per million gallon tank volume.

Careful consideration should be given when selecting the operating sludge age for a SymBio® system as both nitrification and denitrification are performed in the same reactor. During design, the required minimum sludge ages for complete nitrification and denitrification are estimated separately and then they are combined for a single tank operation. Table 6.1 provides the recommended nitrification sludge age and the denitrification rate at various operating temperatures (16).

Once the total, minimum sludge age is calculated based on the loading and the operating temperature, a safety factor is added (20–25%) and the actual operating sludge age is estimated. For example, a municipal wastewater treatment plant operating at 15°C with an influent loading of 220 ppm BOD, 220 ppm total suspended solids (TSS) and 40 ppm total Kjeldahl nitrogen (TKN) and with the effluent requirements of 10 ppm BOD, 10 ppm TSS, 1.0 ppm ammonia-N and 5.0 ppm total nitrogen, will be required to maintain a total sludge age of 11–12 days under the SymBio® mode.

Table 6.1
Sludge age requirements for a simultaneous nitrification and denitrification system

Operating temperature (°C)	Minimum nitrification sludge age (Days)	Denitrification rate (lb nitrate-N/lb VSS-day)
10	10	0.054
15	6	0.066
20	4	0.108
25	4	0.186

Source: (16).

Pochana and Keller (17) indicated that there are three major factors that affect SND, namely, carbon supply, floc size, and dissolved oxygen concentration.

Presence of organic carbon as a food source is essential for denitrification to proceed satisfactorily. According to Henze (18), a chemical oxygen demand to nitrogen ratio (COD:TKN) of 3.5–4.5 g COD per gram of N is necessary for denitrification. In the SymBio® process design, a minimum ratio of 4.0 g BOD per gram of N is recommended to ensure complete denitrification.

Pochana and Keller (17) showed that floc sizes higher than 80 μm allowed significant SND due to development of anoxic regions near the center of the flocs. Kaempfer et al. (7) also demonstrated existence of anoxic regions within 3 mm size flocs using microprobes and suggested that reduced shear force from aeration or mixing equipment could create larger flocs and promote SND. In the SymBio® process, a low DO operation results in a lower aeration energy that reduces the shear on the flocs and promotes SND.

One of the major concerns with low DO operations has been the possible negative effect on the nitrification as well as the denitrification rates. It is generally accepted that denitrification can be best achieved in absence of oxygen whereas nitrification requires approximately 2.0 ppm DO to avoid any rate limitations. Stensel (8) estimated that operating with 0.2 ppm DO concentration at 20°C; the rate of nitrification will be less than one fourth of the rate at 2.0 ppm DO. This means the sludge age in a 0.2 ppm DO operation has to be increased by four times to ensure a complete nitrification compared to a 2.0 ppm operation. Similarly the denitrification rate with a 0.2 ppm DO operation was estimated at half of the rate at 0 ppm DO (8). In the SymBio® process, the DO is consistently maintained below 1.0 ppm to promote SND. However, the field experience has not demonstrated a strong negative effect of a low DO operation on the nitrification or the denitrification rates. In most cases, a complete nitrification is achieved with effluent ammonia concentrations well below 1.0 ppm. Further, the degree of denitrification has been 80% or higher in these cases. Several hypotheses are presented to explain these results:

1. Pochana and Keller (17) observed that in a cyclic study where 95% SND was observed with DO levels in the range of 0.2–0.6 ppm, nitrate was not generated in a significant quantity although the ammonia oxidation was completed. This could have been due to a short SND pathway where the ammonia was oxidized to nitrite only and the nitrite was subsequently reduced to nitrogen gas via dissimilative pathway. Hanaki et al. (19) also indicated that nitrite oxidizers (*Nitrobacter*) were strongly inhibited by a low DO operation. Hence, it is possible that in the SND systems like the SymBio® process majority of the ammonia gets converted to nitrite only and the nitrite in turn gets reduced to nitrogen gas. This possibility provides significant advantages as a shorter pathway reduces the aeration requirement for ammonia oxidation and also reduces the organic carbon requirement during denitrification. Further research to confirm this observation is planned.
2. As discussed earlier, conventional BNR systems involve a pre-anoxic tank followed by a nitrification tank. The recycled mixed liquor from the nitrification tank provides a source of nitrate and the raw influent wastewater provides a source of the organic carbon for denitrification in the pre-anoxic step. The pure aerobic environment in the nitrification stage results in repression of denitrifying enzymes, which have to be activated in the pure anoxic environment for denitrification. Lag periods of 40 min to 2 h have been reported for an aerobically grown culture, for example of *P. aeruginosa*, to shift to maximum denitrification activity (19–21). In the SymBio® process,

the biomass is continuously maintained at DO concentration below 1.0 ppm and is never exposed to fully aerobic conditions. This reduces the extent of enzymatic repression for denitrification and results in relatively a shorter lag period and consequently a quicker shift to denitrification.

3. Hanaki et al. (22) reported that a low DO operation resulted in a higher yield of ammonia oxidizers (*Nitrosomonas*) in a pure nitrification system and this compensated for the reduced ammonia oxidation rate observed at low DO. This means that when the organic loading is low, in the SND systems like the SymBio® process, the nitrification rate per unit biomass may be lower at low DO but the autotrophic yield and hence the quantity of biomass performing the function can be higher compared to a high DO nitrification process.

Another concern with the SymBio® process has been the possibilities of excessive low DO filamentous growth. Further, as SND systems like the SymBio® process usually operate at low food to microorganisms (F:M) ratio, there is a concern for low F:M filamentous bulking. Some of the indicative filamentous microorganisms are Type 1701, *S. natans, H. hydrossis, M. parvicella, Nocardia* sp. Types 021N, 0041, 0675, 0092, 0581, 0961, 0803 (23). However, a study performed at Olympus Terrace wastewater treatment plant in Washington demonstrated that the SymBio® operation did not result in any excessive bulking of the sludge (24). Further, as described in the case study discussing the Rochelle, IL installation in the next section, improvement in the sludge settling has been observed in some cases. One of the reasons could be a selector effect due to the anoxic environment created in the sludge floc because of the low DO operation. This may help in suppressing excessive filamentous growth. Further research to confirm this observation is planned.

In practice, to avoid bulking issues during operation, a small polishing step is included after the SymBio® reactor. The polishing step includes an aerobic tank operating at relatively higher DO (1–2 ppm). The hydraulic retention time is usually 1–2 h. This step helps in reducing any bulking pressure in the system. Further, it polishes off any remaining BOD or ammonia and introduces positive DO into the treated water before the discharge into the secondary clarifiers. Addition of a small pre-anoxic selector basin upstream to the SymBio® basin is also planned to create the necessary F:M gradient and incorporate microbial selection against filamentous organisms.

5. CASE STUDIES

The SymBio® process has been used in more than 30 installations worldwide. Measurements involving NADH, organic loading, respiration, or DO concentration have been used to estimate the oxygen demand of the biomass and to control the air supply to maintain SND in these plants. Some operational data is presented here based on the experience with the SymBio® process in the USA.

5.1. Big Bear, CA

Big Bear Area Regional Wastewater Agency (BBARWA) in California operates an oxidation ditch plant with a design flow of 3.2 MGD. There are three oxidation ditches but only two are normally used in a parallel operation. Each ditch has a volume of 1.6 million gallons and uses brush aerators (total 180 hp installed aeration capacity in each ditch) for aeration and mixing. The plant was designed as a conventional nitrification plant based on

ammonia removal requirements in the past. As the effluent requirement changed to 10.0 ppm total inorganic nitrogen (TIN = ammonia-N + nitrite-N + nitrate-N), the plant opted for the SymBio® technology to introduce simultaneous denitrification within the ditches. NADH Jump control strategy has been used in all ditches. Initially, during a 3-month trial period in the summer of 2000, the brush aerators were controlled in On/Off mode based on NADH profile. The modified operation generated effluent TIN values below 2.0 ppm while the ammonia concentrations were maintained below 0.5 ppm. This clearly indicated that complete nitrification was maintained while simultaneous denitrification was introduced using the SymBio® technology. Comparison of the operating hours of brush aerators in SymBio® mode versus the previous nitrification mode demonstrated energy savings in excess of 30% due to the effective aeration control by the NADH monitoring system for a low DO operation.

Subsequently, in March 2001, VFDs were installed on the brush aerators for better aeration control. NADH was monitored in the biomass and the speed of the aerators was regulated to match the oxygen demand. Effluent nitrogen concentration results from this automatic operation in 2001–2003 are provided in Fig. 6.10 and Fig. 6.11. Table 6.2 provides the average plant effluent results during this period. The capital cost savings for the BNR upgrade using the SymBio® process were in excess of $500,000 as separate anoxic tanks with mixed liquor recycle were not required.

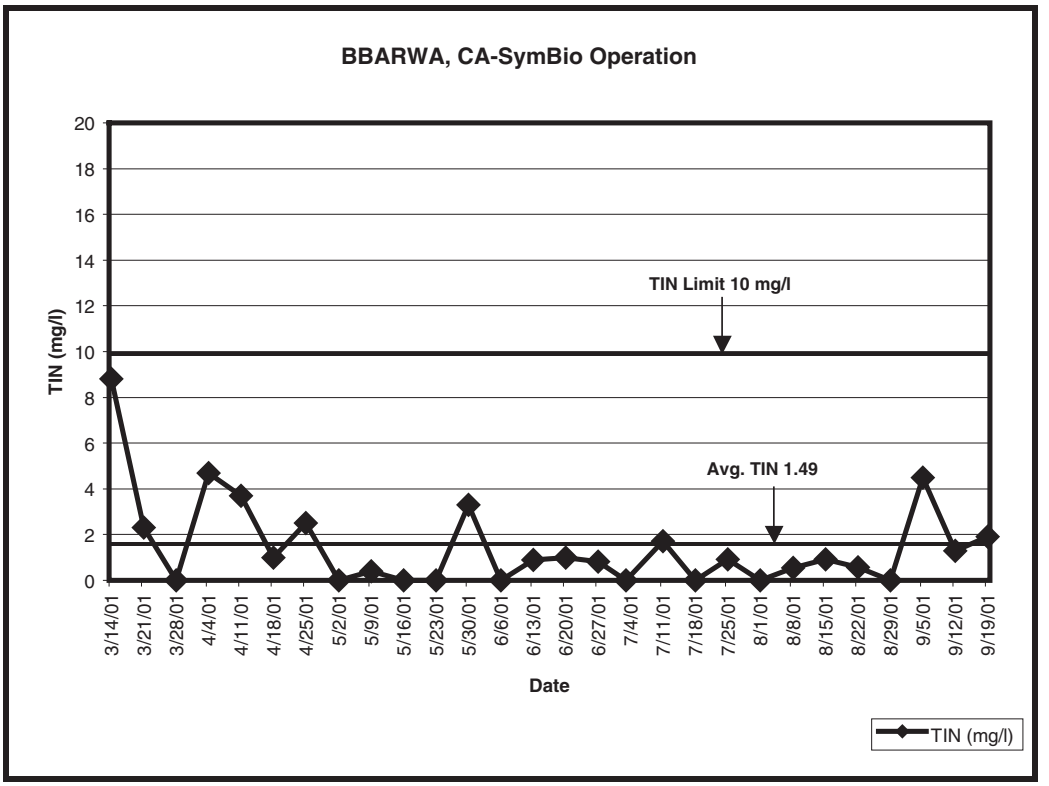

Fig. 6.10. Big Bear, CA-Effluent TIN results.

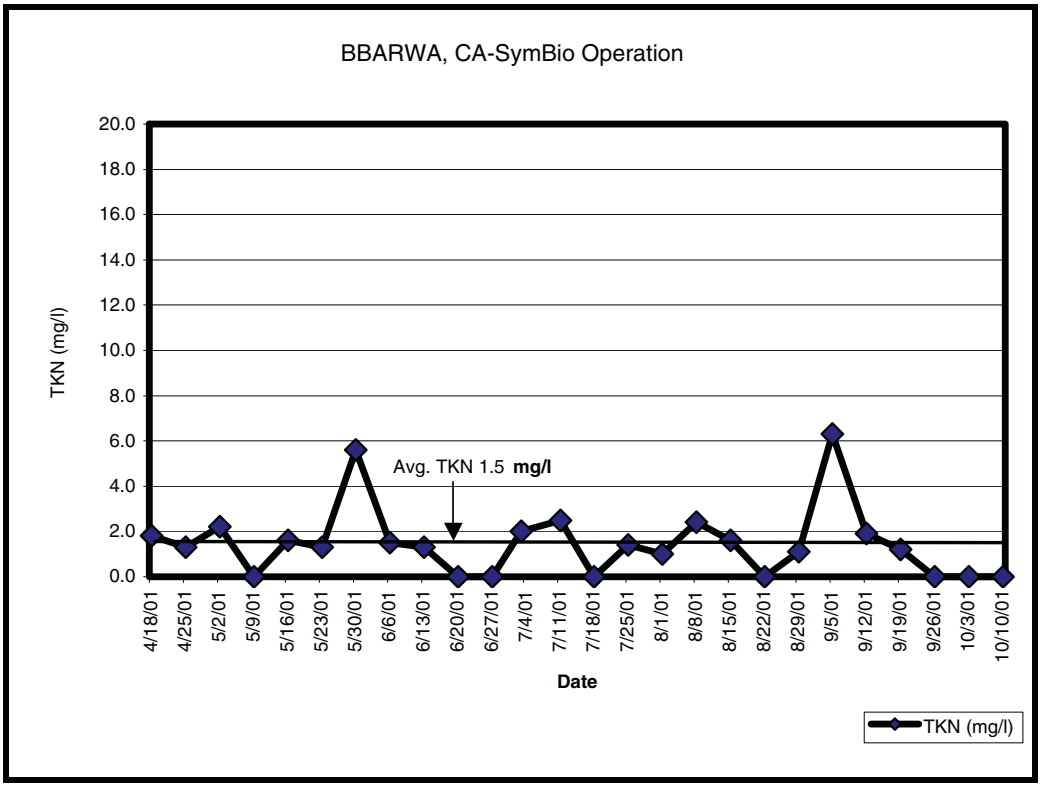

Fig. 6.11. Big Bear, CA-Effluent TKN results.

Table 6.2
Big Bear, CA-Effluent results with the simultaneous nitrification and denitrification system—(2001–2003)

BEFORE SymBio		Location	Flow (MGD)	BOD (ppm)	TSS (ppm)	Ammonia-N (ppm)	Nitrate-N (ppm)	TIN (ppm)
	1999							
		Influent	2.12	193	219			
		Effluent	2.12	5	14	1.20	8.50	9.70
AFTER SymBio	2001–2002							
		Influent	2.72	275	371			
		Effluent	2.72	6	6	0.6	2.27	2.85
	2002–2003							
		Influent	2.09	286	287			
		Effluent	2.09	8	7	1.43	1.49	3.3

BBARWA has won the following awards due to their success with the SymBio® operation:

- Innovation Award (2000) from the California Association of Sanitation Agencies (CASA)
- Research Achievement Award (2001) from the California Water Environment Association (CWEA)

5.2. *Perris, CA*

Eastern Municipal Water District operates the 7.5 MGD, 5-stage Bardenpho plant at Perris, CA. The five stages of the Bardenpho process at this facility are:

- Pre-anaerobic stage
- Pre-anoxic stage
- Nitrification stage (oxidation ditch)
- Post-anoxic stage
- Post-aeration stage

The combination of the pre-anaerobic stage and the aerobic, nitrification stage results in a luxury (excessive) phosphorous uptake by the biomass and this phosphorous, stored inside the bacterial cells, is removed from the system when the excess sludge is wasted. The pre-anoxic and the post-anoxic stages are designed to promote denitrification. So, the 5-stage Bardenpho system is a complete BNR system for the removal of both nitrogen and phosphorous. The wastewater treatment plant at Perris was achieving its effluent requirement effectively. However, the city decided to introduce the SymBio® technology in the nitrification ditch to convert it to a SND system. The objective was energy savings due to a low DO operation in the nitrification ditch coupled with a higher nitrogen removal efficiency. In the summer of 2001, a 3-month trial was initiated with the nitrification stage converted to SymBio® SND system and the results of the trial are presented in Table 6.3.

Table 6.3
Perris, CA-Effluent results with the simultaneous nitrification and denitrification system—Summer 2001

Parameter	Value
Plant Flow	7.05 MGD
Secondary Bardenpho Effluent Ammonia-N	1.30 ppm
Secondary Bardenpho Effluent Nitrate-N	0.36 ppm
Secondary Bardenpho Effluent Orthophosphate-P	0.89 ppm
Tertiary Bardenpho Effluent CBOD	2.02 ppm
Tertiary Bardenpho Effluent TSS	<3.0 *ppm*
Tertiary Bardenpho Effluent Ammonia-N	0.57 ppm
Energy consumption before SymBio™ operation	102.40 KW/MG water treated
Energy consumption during SymBio™ operation	76.19 KW/MG water treated
Energy savings due to SymBio™ operation	25.50%

As the data indicated, running under the SymBio® mode the plant maintained high degrees of nitrification and denitrification. The average secondary effluent nitrate concentration and the tertiary effluent ammonia concentration were below 1.0 ppm. Further, the low DO operation due to NADH control resulted in energy savings in excess of 25% compared to the previous operation. Subsequently, the surface aerators in the nitrification ditch were installed with VFDs and the SymBio® system was permanently installed to control the speed of the aerators.

5.3. Rochelle, IL

This 4.87 MGD plant, operated by Rochelle Municipal Utilities, utilizes a parallel operation between four plug flow reactors, each using a two-pass system. Only two reactors are normally used at a given time. Fine bubble diffusers coupled with centrifugal, multi-stage blowers are used for aeration. The plant treats a combination of industrial (food processing) and domestic wastewater. It is required to perform nitrification only and is not required to denitrify at this point. However, the city decided to install the SymBio® system in 2001 at this facility to maximize the energy savings and to use the NADH measurements for monitoring the organic loading fluctuations from the industry. Further, a denitrification requirement is expected in the future. Because of the industrial contribution, the influent TKN loading is relatively high in a range of 50–60 ppm and correspondingly the effluent nitrates have been high. The installation of the SymBio® system has resulted in simultaneous denitrification (more than 70%) in the plug flow reactors and the overall plant effluent results for the 2001–2003 operation are shown in Table 6.4.

One of the benefits with the SymBio® operation for this facility has been the improvement in the sludge settling. As Table 6.4 indicates, the sludge volume index (SVI) based on 30-min settling tests has been maintained at around 130. No filamentous bulking has been observed. This is important, as one of the concerns with low DO operations has been excessive growth of low DO filamentous organisms.

Table 6.4
Rochelle, IL-Effluent results with the simultaneous nitrification and denitrification system—(2001–2003)

2001–2002	Flow (MGD)	BOD (ppm)	TSS (ppm)	Ammonia-N (ppm)	TKN (ppm)	Nitrate-N (ppm)
Influent	3.04	252.00	148.00		73.00	
Effluent	3.04	1.58	1.01	0.26		17.92
2002–2003	Flow (MGD)	BOD (ppm)	TSS (ppm)	Ammonia-N (ppm)	TKN (ppm)	Nitrate-N (ppm)
Influent	2.08	285.00	166.00		42	
Effluent	2.08	3.15	2.8	0.64	3	8.97
Average 2001–2003		MLSS (ppm) 2184	SVI 131	Sludge Age (day) 15		

6. CONCLUSION

Simultaneous nitrification and denitrification is an attractive option for design engineers and scientists as it may offer significant advantages compared to conventional processes with separate nitrification and denitrification reactors. The SymBio® process for simultaneous nitrification and denitrification offers a relatively easy retrofit option to existing nitrification facilities for total nitrogen removal. It can be applied to various flow configurations such as complete mix systems, plug flow systems, oxidation ditches, conventional BNR systems with multi-stage operation, and sequential batch reactors (SBR). The measurement of NADH fluorescence provides an effective tool for monitoring the biological activity, which in turn is used for a precise control over the aeration in the SymBio® process. The NADH proportional control strategy or the NADH jump control strategy can be used with the SymBio® process. The SymBio® operation results in significant energy savings combined with higher nitrogen removal efficiencies.

Since NADH is an intracellular coenzyme, a loss of NADH activity indicated by a loss of the fluorescence signal is usually an indication of influent toxicity in the system. Further, as the NADH is measured by its fluorescence (optical measurement), the NADH sensor has less maintenance issues compared to the conventional measurement techniques involving wet chemistry. As the aeration control steps are taken based on relative changes in the NADH levels under the NADH jump control strategy, the absolute value of NADH in the biomass has no effect over the overall control scheme. This eliminates any influence of external parameters, like MLSS and temperature, on the process performance. It also eliminates the need for calibration of the NADH sensor.

New technologies for nitrification denitrification processes in wastewater treatment involve the use of membrane bioreactors (MBR). The feasibility studies and practical applications of these new MBR technologies are reported elsewhere (25, 26). The SymBio® process control concept has already been combined with MBR technology in the US wastewater treatment industry.

The SymBio® process is a property of BioBalance A/S, Denmark and is protected in the USA under patents 5,506,096, 5,557,415, 5,700,370 and 5,906,746. Enviroquip, Inc. of Austin, Texas has the exclusive rights to offer the SymBio® process in the USA.

NOMENCLATURE

BNR = biological nutrient removal
CO_2 = carbon dioxide
NH_4^+-N = ammonium-nitrogen
O_2 = oxygen
NO_3^--N = nitrate-nitrogen
$CaCO_3$ = calcium carbonate
BOD = biochemical oxygen demand
H_2CO_3 = carbonic acid
MLE = modified Ludzack-Ettinger process
UCT = University of Cape Town process

A^2O = anaerobic/anoxic/oxic process
SND = simultaneous nitrification and denitrification
DO = dissolved oxygen
NADH = nicotinamide adenine dinucleotide
ATP = adenosine tri phosphate
MLSS = mixed liquor suspended solids
PLC = programmable logic controller
VFD = variable frequency drive
TKN = total Kjeldahl nitrogen
F:M = food to microorganism ratio
TIN = total inorganic nitrogen

REFERENCES

1. National Academy of Science, *Nitrates: An Environmental Assessment*, Panel on Nitrates of the Coordinating Committee for Scientific and Technical Assessment of Environmental Pollutants, Washington, DC (1978).
2. P. Norgaard, K. Helmo, and E. Sorensen, Renseproces til kvcestoffjernelse styret af NADH, *Vand og Jord*., 3, 125–129 (1996).
3. C. W. Randall, J. L. Barnard, and H. D. Stensel (eds.), Principals of Biological Nutrient Removal, *Design and Retrofit of Wastewater Treatment Plants for Biological Nutrient Removal*, pp. 25–84, Technical Publishing Company. Inc., Lancaster, PA (1992).
4. US Environmental Protection Agency, *Process Design Manual for Nitrogen Control*, Office of Technology Transfer, Washington, DC (1975).
5. M. Wagner, G. Rath, H. P. Koops, J. Flood, and R. Amann, In situ analysis of nitrifying bacteria in sewage treatment plants, *Water Science and Technology*, 34, 237–244 (1996).
6. W. J. Payne, *Denitrification,* Wiley, New York (1981).
7. H. Kaempfer, G. Daigger, and C. Adams, *Characterization of the Floc Micro-Environment in Dispersed Growth Systems*, Technical paper presented at WEFTEC 2000, Anaheim, CA (2000).
8. H. D. Stensel, *Biological Nutrient Removal: Merging Engineering Innovation and Science*, Technical paper presented at WEFTEC 2001, Atlanta, GA (2001).
9. L. A. Robertson et al., Simultaneous nitrification and denitrification in aerobic chemostat cultures of thiosphaera pantotropha, *Applied and Environmental Microbiology*, 54, 2812 (1988).
10. K. J. Davies, D. Lloyd, and L. Boddy, The effect of oxygen on denitrification in Paracoccus denitrificans, *Journal of General Micobiology*, 135, 2445 (1989).
11. H. K. Trivedi and N. Heinen, *Simultaneous Nitrification/Denitrification by Monitoring NADH Fluorescence in Activated Sludge*, Technical paper presented at WEFTEC 2000, Anaheim, CA (2000).
12. H. K. Trivedi, (1995). *Monitoring of Microbial Denitrification and Dissimilative Nitrate Ammonification Using an On-line Fluorescence Technique*, Graduate thesis, he University of Akron, OH.
13. T. D. Brock and M. T. Madigan, *Biology of Microorganisms*, 5th ed., Prentice Hall, Englewood Cliffs, NJ (1988).
14. H. K. Trivedi, and T. Weir, *Simultaneous Nitrification and Denitrification Using NADH Fluorescence*, Technical paper presented at the Industrial Wastes Technical and Regulatory Conference, Charleston, SC (2001).
15. Metcalf and Eddy, *Wastewater Engineering; Treatment, Disposal, Reuse*, McGraw-Hill, New York (1991).

16. BioBalance A/S, *Engineering Manual*, Denmark (1999).
17. K. Pochana, and J. Keller, Study of factors affecting simultaneous nitrification and denitrification (SND), *Water Science and Technology*, 39, 61–68 (1999).
18. S. H. Isaacs, and M. Henze, Controlled carbon source addition to an alternating nitrification-denitrification wastewater treatment process including biological P removal, *Water Research*, 29, 77–89 (1994).
19. L. K. Ju and H. K. Trivedi, Monitoring of denitrification by Pseudomonas aeruginosa using an on-line fluorescence technique, *Biotechnol. Tech*., 6, 549–554 (1992).
20. L. M. Nelson and R. Knowles, Effect of oxygen and nitrate on nitrogen fixation and denitrification by *Azospirillum brasilense* grown in continuous culture, *Canadian Jounal of Microbiology*, 24, 1395 (1978).
21. W. J. Payne and P. S. Riley, (1969). Suppression by nitrate of Enzymatic reduction of nitric oxide, *Proceedings of the Society for Experimental Biology and Medicine*, 132, 258.
22. K. Hanaki, C. Wantawin, and S. Ohgaki, Nitrification at low levels of dissolved oxygen with and without organic loading in a suspended-growth reactor, *Water Research*, 24, 297–302 (1990).
23. D. Jenkins, *et al*., Manual on causes and control of activated sludge bulking and foaming, *Water Res. Comm*., Pretoria, South Africa (1984).
24. R. W. Samstag, H. D. Stensel, and H. K. Trivedi, *Full-scale Test and Simulation of Simultaneous Nitrification and Denitrification in an Oxidation Ditch Plant*, Technical paper presented at the 2001 PNPCA conference, Sunriver, OR (2001).
25. Editor *Feasibility of New Technologies for Nitrification Denitrification Processes in Wastewater Treatment*. http://allserv.rug.ac.be/-pboeckx/reactor.htm. August (2003).
26. L. K. Wang and R. Menon, Membrane bioreactors. In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.) Humana Press, Totowa, NJ (2009).

7
Single-Sludge Biological Systems for Nutrients Removal

Lawrence K. Wang and Nazih K. Shammas

CONTENTS

Abstract In a conventional activated-sludge process, bio-oxidation, nitrification, and denitrification reactions occur in three separate bioreactors connected in series. Each bioreactor has its own type of micro-organisms (i.e., activated sludge), and each bioreactor has its own clarifier for micro-organisms–water separation. In a single sludge biological system, the mixed micro-organisms are used throughout the bioreactor, which is divided into aerobic and anoxic zones for nutrient removal. This chapter introduces the classification, stoichiometric principles, kinetic considerations, and system design of various single sludge biological systems. Specifically, the multistage single anoxic zone system, the multistage multiple anoxic zone system, and the multiphase cyclical aeration system are discussed in detail. Other biological

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_7

systems covered in this chapter are: endogenous nitrate respiration, aerobic sludge synthesis, anoxic biosolids synthesis, and compartmentalized aeration tanks.

Key Words Anoxic biosolids synthesis • anoxic zone • bio-oxidation • compartmentalized aeration • denitrification • endogenous nitrate respiration • ENR • multiphase cyclical aeration • nutrients removal • nitrogen removal • nitrification • single sludge biological system.

1. INTRODUCTION

Single-sludge nitrification–denitrification processes were first developed and applied in the 1960s. Since then, these processes have gained popularity, particularly in small- to medium-sized plants. Driving factors include less-critical energy and onsite tankage considerations, an increase in the general understanding of basic process principles, marketing efforts by companies promoting proprietary single-sludge systems and a perception that such systems offer potential cost advantages over multiple-sludge nitrogen removal processes and systems with separate-stage denitrification (1).

Single-sludge systems for nitrogen removal basically combine carbonaceous removal, ammonia oxidation, and nitrate reduction within the same process, using modified versions of the activated sludge process with a single sedimentation step for separation of the biosolids. As the enforcement of effluent nitrogen limits became more prevalent in various parts of the country, increased efforts were made to develop new or modified versions of the single-sludge process. As a result, there is now a wide variety of system configurations from which to choose. Single-sludge systems have been developed with various combinations of single or multiple anoxic zones (2–10), oxidation ditches (11–15), sequencing batch reactors (16–24), and cyclical aeration systems (25–28).

Single-sludge systems are available with a variety of design layouts, reactor configurations, inlet feed arrangements, compartmentalization or baffling, mixing processes, return biosolids requirements, internal recycle patterns, aeration processes, integrated phosphorus removal techniques, performance capabilities, process control requirements, and miscellaneous support approaches and controls. This chapter is intended to assist the reader in screening, evaluating, and/or selecting, if appropriate, a single-sludge system. The chapter also provides information on the types of systems, design considerations, and features of various configurations, support systems, performance capabilities, operational requirements, and other factors to consider in designing new plants, plant expansions, and retrofits of existing plants (29–31) for both nitrogen and phosphorus removals. One year operational data of a model single sludge activated sludge treatment plant in Coxsackie, New York, USA, are presented in this chapter in detail for practical engineers to use as a reference. The Coxsackie plant was designed for both nitrogen and phosphate removal, and has been in successful operation since 1974. More detailed design examples can be found from Chapter 8, Selection and Design of Nitrogen Removal Processes.

Single-sludge systems offer several advantages over multiple-sludge systems or separate-stage systems (1):

1. Without intermediate clarifiers or separate denitrification units, there is a potential cost advantage, if the costs of larger reactor tankage and energy requirements do not exceed these benefits.
2. Space availability.
3. Reduction in alkalinity consumption.
4. Use of wastewater carbon as a carbon source for denitrification in lieu of methanol.
5. Lower oxygen requirements.
6. Single-sludge systems can more readily be used in retrofitting existing activated sludge plants for nitrogen removal, particularly if the plant has excess capacity.

Potential limitations or disadvantages (which are site specific and all or none may apply to a particular situation) to consider compared to separate sludge/stage systems include (1):

1. Greater sensitivity to toxicity or inhibition without a separate upstream biological treatment step.
2. Lower nitrogen removal efficiency.
3. Higher energy usage (compared to separate stage).
4. Larger volumes of reactor tankage.
5. Greater site requirements.

The major factor—in addition to the effluent nitrogen limit—in evaluating and comparing a single-sludge system to other systems is cost comparison in terms of capital outlay and operation and maintenance. It must also be noted that single-sludge systems can be followed by a separate stage for denitrification where more stringent nitrogen limits are imposed. The separate stage may need to be operated during winter only while operating the single-sludge system exclusively for nitrification. During warmer months, the single-sludge system would be used for nitrogen removal without the separate stage, thus eliminating methanol costs.

Phosphorus and phosphates in the wastewater can be removed significantly in a biological treatment plant with or without chemical additions. Although the emphasis of this chapter is on nitrogen removal by single sludge activated sludge processes, this chapter also discusses how phosphorus and phosphates can be removed at the same time in a biological reactor.

2. CLASSIFICATION OF SINGLE-SLUDGE PROCESSES

Single-sludge systems are generally classified according to their flow regime, staging of anoxic and aerobic sequences, or method of aeration. All the classifications and their component processes require nitrification to occur in an aerobic zone or reactor, followed by denitrification. For denitrification to occur, nitrates must be present together with an organic carbon source. Organic carbon can be provided by the endogenous activity of the micro-organisms (i.e., by depleting the cell's mass) or by an exogenous source such as the BOD of the influent waste-water or primary effluent.

To use endogenous activity as the carbon source, plant flow would be conveyed sequentially through a combined BOD removal/nitrification step in an aerobic zone or reactor, and then to the endogenous anoxic zone or reactor to denitrify the nitrates. Alternatively, the influent BOD can be exploited for denitrification by either:

1. Recycling nitrates to an anoxic zone or reactor that precedes the aerobic zone.
2. Operating alternate anoxic/aerobic conditions within a single zone or reactor.
3. Conveying the flow sequentially through alternating anoxic/aerobic zones.

Because denitrification cannot occur without nitrification occurring first, systems are designed and sized to completely nitrify the oxidizable influent TKN (Total Kjeldahl Nitrogen). Thus, conventional parameters such as F/M (Food to micro-organism) ratio, retention time, oxygen transfer rate, and solids retention time (SRT or θ_c) are used in sizing the aeration equipment and tank volume. Denitrification can then be achieved by conveying the oxidized nitrogen in the form of nitrates to an anoxic zone. A summary of the categories and characteristics of the general single-sludge classifications is provided below (1).

1. *Multistage Single Anoxic Zone*. Processes are most commonly configured as suspended growth treatment. Variations in aeration conditions are achieved spatially in different reactors as flow is conveyed through the process train. This process uses one anoxic stage for denitrification and represents one of the simplest configurations for nitrogen removal in a single-sludge system. The most common configuration to achieve denitrification involves recycling nitrified mixed liquor to an antecedent anoxic zone, where exogenous carbon provided by the influent wastewater can be used by the facultative denitrifiers. Nitrates that are not recycled will be discharged to the final clarifier. Examples of single anoxic zone processes include anaerobic/anoxic/oxic (A^2/O), Modified Ludzack-Ettinger (MLE), Virginia Initiative Plant (VIP), and University of Capetown (UCT) processes.
2. *Multistage Anoxic Zones*. This configuration uses more than one anoxic zone. Two anoxic zones are most commonly used. The carbon source for denitrification may be either endogenous or exogenous; however, endogenous denitrification should be preceded by an exogenous denitrification reactor for maximum nitrogen removal. Endogenous denitrification is commonly used to denitrify the nitrates that were not recycled to the antecedent exogenous denitrification reactor.

Exogenous denitrification can be achieved by the following design strategies: (a) recycling nitrified mixed liquor to an antecedent anoxic zone, (b) step-feeding raw wastewater or primary effluent to an anoxic zone containing nitrates, or (c) supplementing the depleted carbon in the nitrified mixed liquor with methanol. For systems that denitrify by employing two exogenous zones with internal recycle and no endogenous zone, the final effluent nitrate concentration is controlled by the recycle rate since the aerobic zone is not followed by another anoxic zone. This process configuration does not achieve effluent TN (total nitrogen) concentrations as low as configurations that have an endogenous anoxic zone following BOD removal/nitrification.

Step-feeding raw wastewater or primary effluent to provide substrate for exogenous denitrification requires a final aeration step to nitrify the ammonia that bypasses the initial BOD removal/nitrification process.

The Bardenpho and Modified UCT processes are examples of dual anoxic zone processes.

1. *Multiphase/Cyclical Aeration*. Cyclical technologies are generally a modification of the activated sludge process. Alternating anoxic/aerobic sequences are achieved in continuous flow reactors or compartments by pulsing the aeration source. The aeration frequency or intensity should be adjusted such that the DO (dissolved oxygen) in the reactor does not exceed 2 mg/L during the aerobic phase. If several alternating reactors or zones are used in series, raw wastewater or primary effluent may be step-fed to those reactors in which wastewater organic carbon has been depleted or is present in rate-limiting concentrations.
2. *Oxidation Ditches*. Oxidation ditches are perhaps the simplest treatment scheme, but are less common in the United States than conventional activated sludge configurations. Wastewater flows

in a continuous circuitous path and aeration is provided at fixed points along the flow path. Anoxic conditions are achieved between the aerators as oxygen is depleted. The hydraulic retention time of an oxidation ditch is generally longer than in multistage systems (11–15).

3. *Sequencing Batch Reactors (SBRs)*. SBR technologies are among the oldest technologies. By pulsing the aeration mechanism on a timed cycle, alternating aerobic and anoxic conditions are achieved on a temporal basis within a single reactor, as opposed to a spatial basis, and all reactions and settling occur in the same reactor (16–24).

This chapter deals with the first three categories. The last two namely oxidation ditches and sequencing batch reactors (SBRs) are discussed in detail in another book *Biological Treatment Processes* (2), which is also in the same book series: *Handbook of Environmental Engineering*.

3. STOICHIOMETRIC AND KINETIC CONSIDERATIONS

In this section, aspects of single-sludge systems for nitrogen removal are discussed to present the important basis and working theory for the processes discussed in this chapter. The discussion is based on data found in the literature, particularly in a report (32) and manual (33) published by the US Environmental Protection Agency (US EPA).

This section represents an attempt to postulate the principles of certain complex biological processes. As in other attempts of this nature, a good measure of idealization and simplification had to be used.

3.1. Routes of Nitrogen Removal in Single-Sludge Systems

Four routes of nitrogen removal seem to be available in single-sludge systems. These routes are (32–36).

3.1.1. Biosolids Synthesis

This may occur either under aerobic or anoxic conditions. The stoichiometric equation describing biosolids (sludge biomass) synthesis is the same for both conditions. In the course of the synthesis reaction, ammoniacal or organically bound nitrogen is removed from the substrate by the biomass and incorporated into new biomass. The energy deficiency of the synthesis reaction is covered by the energy surplus of aerobic or nitrate (substrate) respiration.

3.1.2. Substrate Nitrate Respiration

This is the interaction of nitrates with organic wastewater carbon compounds and is mediated by the biomass. This reaction always accompanies biosolids synthesis under anoxic conditions. Nitrates are used as the hydrogen acceptor, organic wastewater carbon compounds as the hydrogen donor. This is an energy reaction. Nitrogen is released to the atmosphere. During this reaction, biodegradable carbon compounds are present in the liquid phase of the process water.

3.1.3. Endogenous Nitrate Respiration (ENR)

This is the interaction of nitrates with the biosolids themselves under anoxic conditions. Nitrates are used as the hydrogen acceptor; the biomass itself serves as the hydrogen donor. This too is an energy reaction. Nitrogen is released to the atmosphere and ammonia is released

to the process water. During ENR, biodegradable carbon compounds are absent from the liquid phase of the process water. Adsorbed carbon nitrate respiration often accompanies ENR. Because the chemistry of the biomass and of adsorbed carbon compounds is not well known, it is often impossible to differentiate quantitatively between ENR and adsorbed carbon nitrate respiration. The term ENR is, therefore, used to cover both denitrification based on biomass destruction and denitrification based on the use of adsorbed carbon compounds as the hydrogen donor.

3.1.4. Adsorbed Carbon Nitrate Respiration

This is the interaction, under anoxic conditions, of nitrates with organic carbon compounds adsorbed by the biomass and occurring while the liquid phase of the process water is already devoid of biodegradable carbon compounds. The adsorbed carbon compounds are used as the hydrogen donor by the biomass; nitrates serve as the hydrogen acceptor. This too is an energy reaction. Nitrogen is released to the atmosphere. The reaction occurs in conjunction with ENR and is accompanied by a measure of biosolids synthesis. The stoichiometry of this reaction is probably similar to that of substrate nitrate respiration, but not enough data are available to allow the postulation of stoichiometric equations. For reactor design purposes, the kinetic equations describing ENR can be easily adapted to reflect the occurrence of adsorbed carbon nitrate respiration.

3.2. Stoichiometric and Metabolic Principles

This stoichiometric discussion rests on the following basic assumptions (32):

1. The composition of the biomass produced in the process is $C_5H_7O_2N$ (37)
2. The composition of the carbon source in domestic wastewater is $C_{10}H_{19}O_3N$ (38, 39)
3. The energy/synthesis ratio ($f_e:f_s$) for aerobic biological removal of the carbon source is 0.54:1
4. The energy/synthesis ratio ($f_e:f_s$) for anoxic biological removal of the carbon source is 1.86:1
5. The BOD_5/COD ratio of the carbon source is 0.45:1

The $f_e:f_s$ ratio is the ratio of substrate utilized for energy to substrate utilized for synthesis in a reaction of zero SRT, i.e., in a reaction in which no endogenous respiration occurs. The nomenclature $f_e:f_s$ is after McCarty (38).

The carbon source does not include free ammonia in the domestic wastewater. The energy/synthesis ratio for the aerobic process is a rounded version of the ratio stipulated by Porges et al. (37); the $f_e:f_s$ ratio for the anoxic process was estimated on the basis of data reported by Barnard (40).

From the assumptions listed, several conclusions may be readily drawn:

1. The yield factor Y based on BOD_5 (mg MLVSS produced/mg BOD_5 removed) in aerobic substrate utilization is 1.02; Y is 0.46 in terms of COD.
2. During anoxic substrate utilization, Y is 0.55 in terms of BOD_5, 0.25 in terms of COD.
3. One mole of the carbon source in domestic wastewater ($C_{10}H_{19}O_3N$) has a COD of 400 g, a BOD_5 of 180 g, and a TOC of 120 g and weighs 201 g.
4. One mole of the biomass (MLVSS) produced has a COD of 160 g, weighs 113 g and contains 12.4% nitrogen.

5. The stoichiometric equation for biosolids synthesis under aerobic as well as anoxic conditions is:

$$C_{10}H_{19}O_3N + 1.5NH_3 + 2.5CO_2 \rightarrow 2.5C_5H_7O_2N + 3H_2O \quad (1)$$

It should be noted that NH_4^+-N is assumed to be the source of the additional nitrogen needed, even under anoxic conditions.

6. The stoichiometric equation for aerobic (oxygen) respiration in the presence of substrate is:

$$C_{10}H_{19}O_3N + 12.5O_2 \rightarrow 10CO_2 + 8H_2O + NH_3 \quad (2)$$

Usually this reaction is simply referred to as "respiration."

7. The stoichiometric equation for substrate nitrate respiration is:

$$C_{10}H_{19}O_3N + 10NaNO_3 \rightarrow 10CO_2 + 3H_2O + NH_3 + 10NaOH + 5N_2 \quad (3)$$

8. The stoichiometric equation for aerobic removal (synthesis+respiration) of the carbon source reads:

$$C_{10}H_{19}O_3N + 4.375O_2 + 0.625NH_3 \rightarrow 1.875CO_2 + 4.75H_2O + 1.625C_5H_7O_2N \quad (4)$$

9. The stoichiometric equation for anoxic removal (synthesis+respiration) of the carbon source reads:

$$C_{10}H_{19}O_3N + 6.5NaNO_3 \rightarrow 0.125NH_3 + 5.625CO_2 + 0.875C_5H_7O_2N + 6.5NaOH + 3.25N_2 + 3H_2O \quad (5)$$

In Eqs. (3) and (5), $NaNO_3$ stands for all the nitrates present.

Construction of Eqs. (1) to (5) is based on the half-reaction equations originally proposed by McCarty (38). The heuristic strategy of looking on all biological substrate removal processes as composed of a respiration (energy) reaction and a synthesis (biosolids production) reaction was introduced into environmental engineering by Porges et al. (37). Additional details on the subject of this subsection may be found in reference 41.

3.3. Endogenous Nitrate Respiration (ENR)

ENR was introduced to wastewater treatment in 1964 by Wuhrmann of Zurich (42) who inserted an anoxic reactor between the aeration tank and final settler of a conventional activated sludge system. The anoxic treatment tank was equipped with mixing devices to keep the biomass in suspension.

Equations (6) and (7) describe endogenous oxygen respiration (EOR) and ENR, respectively. Equation (6) has been stipulated by Porges et al. (37) and Eq. (7) by Christensen et al. (44). From Eqs. (6) and (7), Eqs. (8) and (9) were derived by considering the reactions between CO_2, NH_3, and NaOH that may be predicted to occur in the process water (for further details, see Ref. 41).

$$C_5H_7O_2N + 5O_2 \rightarrow 5CO_2 + NH_3 + 2H_2O \quad (6)$$

$$C_5H_7O_2N + 4NaNO_3 \rightarrow 5CO_2 + NH_3 + 2N_2 + 4NaOH \quad (7)$$

$$C_5H_7O_2N + 5O_2 \rightarrow 4CO_2 + NH_4HCO_3 + H_2O \quad (8)$$

$$C_5H_7O_2N + 4NaNO_3 + H_2O \rightarrow 4NaHCO_3 + NH_4HCO_3 + 2N_2 \quad (9)$$

The same amount of ammonia nitrogen is released in both the EOR and ENR reactions. In the aerobic reaction, the ammonia nitrogen is immediately nitrified. This is not the case in the anoxic reaction. Extra aerobic reactor space must be provided if it is desired to oxidize the ammonia nitrogen formed during ENR.

The release of ammonia nitrogen by the biomass gives rise to a slight momentary increase in alkalinity. In the anoxic reaction, additional alkalinity is released due to the reduction of nitrates (3.57 mg alkalinity as $CaCO_3$/mg NO_3^--N gasified). This is one-half of the amounts used up in the nitrification of 1 mg of ammonia nitrogen. Once the nitrates are exhausted, the plant operator will be confronted with unneeded reactor space if the anoxic zone is too large. From Eq. (9), the values of Table 7.1 have been abstracted.

Of particular importance with ENR are two stoichiometric relationships:

(a) For 4 mg of NO_3^--N gasified, 1 mg of NH_4^+-N is released back to the process water
(b) Two milligram of biomass are destroyed for each mg of NO_3^--N reduced; 2.69 mg of biomass are destroyed for each mg of N removed

Table 7.1 concerns a single-stage reaction. All nitrogen removed in this reaction from the mixed liquor is removed by gasification of NO_3^--N. But the amount of NO_3^--N gasified is larger than the net amount of N removed from the liquid phase of the process water. The difference is the ammoniacal N released by the biomass that was destroyed during ENR.

If it is desired to remove the residual ammonia nitrogen generated during ENR, a cascade approach would have to be followed if the basic scheme suggested by Wuhrmann is used. The

Table 7.1
Endogenous nitrate respiration stoichiometric relationships

1 mg of NO_3^-–N gasified	2 mg of biomass destroyed
1 mg of NO_3^-–N gasified	0.25 mg of NH_4^+–N released to the liquid phase of the process water due to biomass destruction
1 mg of NO_3^-–N gasified	1.07 mg of biomass carbon destroyed
1 mg of NO_3^-–N gasified	4.46 mg of total alkalinity produced: 3.57 mg due to reduction of NO_3^-–N, 0.89 mg due to biomass destruction
1 mg of NO_3^-–N gasified	0.75 mg of N removed from the liquid phase of the process water: 1 mg of NO_3^-–N is gasified, but 0.25 mg of NH_4^+–N is added
1 mg of N removed[a]	2.69 mg of biomass destroyed
1 mg of N removed[a]	1.43 mg of biomass carbon destroyed
1 mg of N removed[a]	1.33 mg of NO_3^-–N gasified

Source: US EPA.
[a] From the liquid phase of the process water.
Of particular importance with ENR are two stoichiometric relationships:
(a) For 4 mg of NO_3^-–N gasified, 1 mg of NH_4^+–N is released back to the process water.
(b) Two mg of biomass are destroyed for each mg of NO_3^-–N reduced; 2.69 mg of biomass are destroyed for each mg of N removed.

NH_4^+-N released in the first ENR zone would be nitrified in a following aerobic zone. The nitrate nitrogen generated there would be reduced in a second ENR zone. This would result in a second NH_4^+-N residual, amounting to approximately one quarter of the first residual. Repeated application of the cascade approach would result in an overall process in which each 1 mg of nitrogen removed from the liquid phase of the process water would equal to 1 mg of NO_3^--N gasified.

3.4. Nitrogen Removal by ENR and Aerobic Sludge Synthesis

The effect of increasing rates of biomass destruction via ENR is shown in Table 7.2. It is assumed that 100 mg of BOD_5 are removed and that the initial biosolids volatility is 80%. The first column represents a theoretical process with no biomass destruction. Nitrogen removal is 12 mg, due only to biosolids synthesis. The 80% volatility stays unchanged. In the second column, 42% biomass destruction occurs, resulting in a final biosolids volatility of 70%. A part of this biomass destruction is assumed to be owing to EOR, with no benefit for nitrogen removal. Ten percent of the initial biomass of 100 mg, or 10 mg, is assumed to have been lost in this way. The nitrogen removal in this process is 19 mg. If EOR had been used for biomass destruction, the removal would have been 7 mg. The third column represents a process with 63% biomass destruction. Ten percent of the original biomass of 100 mg is lost again in EOR. The final biosolids volatility is 60%; nitrogen removal is 24 mg. If EOR had been used throughout, nitrogen removal would have been 4 mg. In Table 7.2, it was assumed that 2.69 mg of biomass were destroyed for each milligram of N removed.

The assumed 80% value for initial volatility is probably conservative. The actual value may be somewhat higher; 82% volatility, for instance, was reported for the Newtown Creek return biosolids at SRT of 3 days with pure oxygen operation (44). On the other hand, initial volatility might be much lower than 80% due to metal salt addition for phosphorus removal. The final volatility of 60% is not unattainable. A manufacturer of aeration equipment for oxidation ditches (45) lists 55% volatility as attainable and as characteristic of "completely mineralized" biosolids that do not require further stabilization. Efficient nitrogen removal by

Table 7.2
Stoichiometric projection of nitrogen removal via biosolids synthesis and endogenous Nitrate Respiration[a]

Final biosolids volatility	(%)	80	70	60
Biomass destruction	(%)	–	42	63
Initial biomass	(mg)	100	100	100
Biomass used in ENR	(mg)	–	32	53
Biomass lost in EOR	(mg)	–	10	10
Biomass final	(mg)	100	58	37
N removed in biosolids	(mg)	12	7	4
N removed via ENR	(mg)	–	12	20
N removed total	(mg)	12	19	24

Source: US EPA.
[a] Based on 100 mg of BOD_5 removed and 80% initial sludge volatility.

ENR goes hand in hand with substantial destruction of the volatile biomass generated during BOD removal. This should favorably affect the cost of biosolids disposal.

The relationship between biomass destruction and change in biosolids volatility may be expressed by Eqs. (10) and (11):

$$v = \frac{v_0(1-b)}{1-bv_0} \tag{10}$$

$$b = \frac{v_0 - v}{v_0(1-v)} \tag{11}$$

where

v = final fractional biosolids volatility, decimal fraction or %
v_0 = initial fractional biosolids volatility, decimal fraction or %
b = fractional rate of biomass destruction, decimal fraction or %

The validity of these equations may be recognized by considering for instance that the inert mass of biosolids containing 60% VSS was originally associated with 160 mg of VSS if v_0 was 0.8, as illustrated in the following example:

In the course of an endogenous respiration procedure, 200 mg of dried biosolids of 80% initial volatility ($v_0 = 0.80$) are reduced by 100 mg. Assuming that the entire weight reduction was with respect to volatile solids, the composition of the solids at the end of the procedure will be 40 mg fixed and 60 mg volatile ($v = 0.60$). The destruction of the volatile biomass amounts to 100 mg. This is (100/160) $100 = 62.5\%$ biomass destruction. Using Eq. (11) furnishes the same result:

$$b = \frac{0.80 - 0.60}{0.80(1-0.60)} = 0.625 \text{ or } 62.5\%$$

In the operation of systems using ENR and aerobic biosolids synthesis for N removal, two conditions must obviously be avoided or minimized: anoxic residence time in the absence of nitrates and loss of wastewater carbon in EOR. To minimize the occurrence of these conditions, it appears advisable to use a compartmentalized reactor, equipped with aeration and mixing equipment in such a way that the bulk of the compartments may easily be switched over from the aerobic to the anoxic condition and vice versa.

Two basic aeration patterns are conceivable, the block approach and the alternating zones approach, as shown in Fig. 7.1. A compartmentalized reactor equipped with dual equipment (aeration and mixing) in most compartments would allow the use of either approach. In both approaches, the last cell should be an aerobic cell to nitrify ammonia nitrogen released in preceding cells and to strip gaseous nitrogen clinging to the floc.

If a reasonable number of cells are provided under the block approach, the daily aeration pattern could be modified in response to changes in load and reaction rates. When using the alternating zones approach, such adjustments might not be necessary.

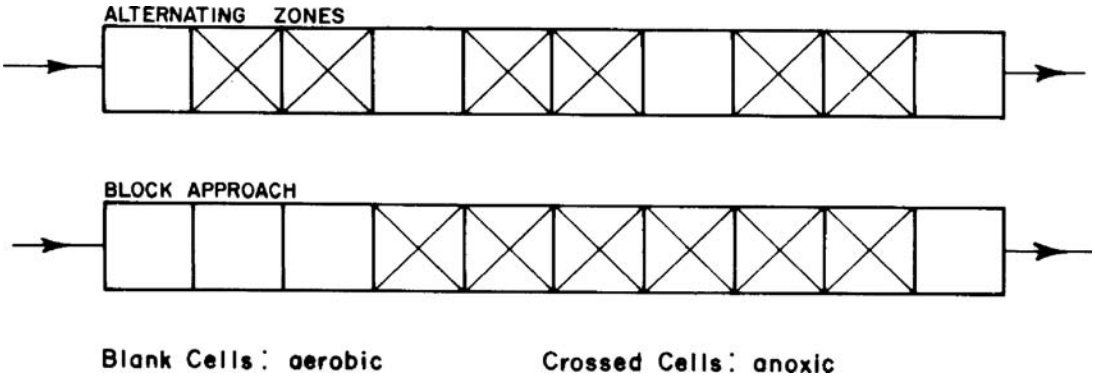

Fig. 7.1. Aeration patterns in compartmentalized reactors. (*Source*: US EPA).

Table 7.3
Substrate nitrate respiration and anoxic synthesis

Reduction of 1 mg of NO_3^-–N is related to		
Species	Action	mg
TOC	Removed	1.32
BOD_5	Removed	1.98
COD	Removed	4.40
Biomass	Produced	1.09
TKN	Metabolized	0.13
NH_4^+–N	Released	0.02
Alkalinity ($CaCO_3$)	Released	3.57

Source: US EPA.

3.5. Nitrogen Removal by Substrate Nitrate Respiration and Anoxic Biosolids Synthesis

The stoichiometric equation for anoxic synthesis and substrate nitrate respiration combined in the ratio of 0.54:1 was listed as Eq. (5). From that equation, Table 7.3 was prepared.

Approximately 2 mg of BOD_5 are needed for the gasification of 1 mg of $N0_3^-$-N. This process of gasification also produces 1.09 mg of biomass, metabolizing thereby 0.13 mg of TKN. Whether or not NH_4^+-N is released in the reaction depends on the nitrogen content of the organic carbon source. Assuming that the organic carbon source has the composition $C_{10}H_{19}O_3N$, the net release of N to the process water (in the form of NH_4^+-N) is 0.02 mg per mg of $N0_3^-$-N gasified. The biomass generated during substrate nitrate respiration and anoxic synthesis, obviously, could be used for nitrogen removal via an ENR anoxic reactor further downstream.

Substrate nitrate respiration is somewhat more difficult to implement in a single-sludge system than endogenous nitrate respiration. It becomes necessary to arrange for contact

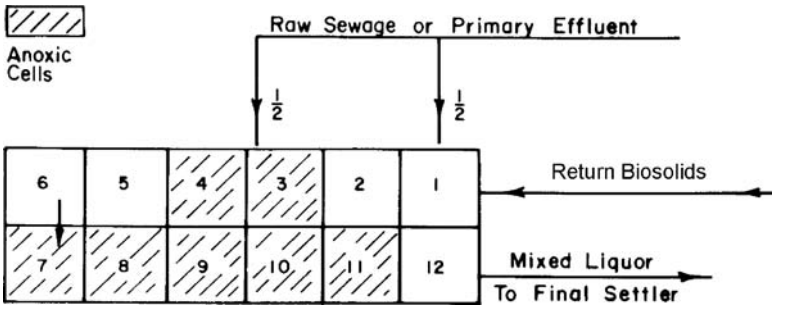

Fig. 7.2. Two-step feeding of compartmentalized reactor. (*Source*: US EPA).

between nitrate nitrogen and untreated process water. But nitrates only become available after the process water has received a considerable measure of treatment. However, the carbon source in the substrate is removed from the process water after a very short length of treatment, this phenomenon being the basis of the contact stabilization process. There seems to exist, basically, three ways to overcome these apparently contradictory requirements:

1. *By step feeding* the untreated process water to the reactor as illustrated in Fig. 7.2. Only a portion of the carbon source in the wastewater is utilized for substrate nitrate respiration in this process, which is discussed at some length in Ref. (41). The process water is divided into two or more portions. The first portion is fed, together with the return biosolids, into the first aerobic cell of a compartmentalized aeration tank. The remaining portions are fed into anoxic cells further downstream. Interaction between the organic carbon sources in these streams with nitrates generated in preceding aerobic cells brings about substrate nitrate respiration.
2. *By mixed liquor recirculation*, as suggested by Ludzak and Ettinger in 1962 (46). This procedure, which is discussed in a later section, is now used in the first two reactors of the patented Bardenpho process (40, 47). The stoichiometry of this process is also discussed in detail in Ref. (41).
3. *By pulsating aeration*. The untreated process water is introduced into a basin that is alternately aerated and mixed anoxically. At the end of the aerobic period, the nitrogen of the process water will be mostly in the nitrate form. During the ensuing anoxic period, substrate nitrate respiration will occur via interaction between the inflowing untreated process water and these nitrates. However, the inflowing ammonia will remain nearly intact during the anoxic period. Nitrifiers are inactive at very low dissolved oxygen. One way to prevent substantial bleedthrough of NH_4^+-N would be to place a nitrification reactor after the pulsating aeration basin. To obtain an effluent very low in NO_3^--N, some ENR will have to be provided as an additional step. For domestic wastewater, ENR does not occur in the pulsating aeration basin itself. During the anoxic periods, the available NO_3^--N is completely utilized by the organic wastewater carbon source present. At the end of the anoxic period, the nitrates will be exhausted but some organic wastewater carbon source will still be available. In order for ENR to occur, NO_3^--N must be present but substrate must be absent. Several implementations of the pulsating aeration concept are discussed in the US EPA Manual (33).

One cannot compare the nitrogen removal capacity (mg N removable/mg BOD_5 removed) of ENR with the removal capacity of substrate nitrate respiration per se. Removal capacities are comparable only in the framework of projected flow sheets. According to the third column

of Table 7.2, the limit of the nitrogen removal capacity of ENR plus aerobic biosolids synthesis is not far above 0.24 mg N/mg BOD removed. Reference (40) indicates that the Bardenpho flow sheet has approximately twice this nitrogen removal capacity. Under that flow sheet, the bulk of the nitrogen is removed via substrate nitrate respiration.

3.6. Design Alternatives for Compartmentalized Aeration Tanks

The reactor should be designed as a compartmentalized tank to produce a near plugflow hydraulic regime and to allow for flexibility in the aeration pattern (32). All biological reactions are far from being zero order reactions under low substrate or hydrogen acceptor (NO_2^- and NO_3^- nitrogen) concentration, and such low concentrations are often prescribed under today's more stringent effluent requirements. The plugflow hydraulic regime is obviously superior to the completely-mixed hydraulic regime for all reactions of an order greater than zero. Aeration pattern flexibility is important to achieve full reactor utilization in all seasons.

The block arrangement of aerobic-anoxic cells is one of the two basic patterns of aeration cells that seem to be available, as illustrated previously in Fig. 7.1. The other arrangement is the alternating zones approach. Flexible mechanical design will allow the use of either approach in the same reactor. However, the alternating zones approach, if used in a reactor of generous overall design, might make it unnecessary to adjust the aeration pattern to seasonal changes of operating conditions.

A recommended cell size is $30 \times 30 \times 30$ ft ($9.1 \times 9.1 \times 9.1$ m), equivalent to a cell volume of 202,000 gal (754 m^3) (32). For gentle mixing, a power input of 10 W/m^3 is required. The inlet and outlet ports have to be near the bottom because the solids are not uniformly dispersed throughout the tank cells. Dissolved oxygen control equipment should be provided for any cell where aeration equipment is installed. An excessive dissolved oxygen concentration, over 3 mg/L, will impair the efficiency of the following anoxic cell.

A large biosolids recycle rate will depress the substrate gradient in a compartmentalized reactor; this is kinetically disadvantageous because the average substrate concentration in a plugflow reactor determines its removal rate. On the other hand, a high recycle rate often makes management of the solids in the final settler easier.

Exclusive of the aeration system itself, a decision must be made whether to provide either primary settling or flow equalization or both. To cope with load fluctuations, the designer has the choice, in many cases, between a larger aeration tank and flow equalization. In a plugflow aerobic-anoxic reactor, extension of flow-through time up to a certain point will not cause the release of nutrients nor deterioration of the effluent via deflocculation, assuming that phosphorus is controlled by metal salt addition; it will, however, result in decreasing biosolids production. The designer should also be cognizant that flow equalization imposes a certain measure of complexity on the operation and maintenance of a plant.

It is, therefore, recommended that increasing the reactor size be given priority consideration unless there are other factors beside the secondary system that can only be resolved by flow equalization. If flow equalization is not included in the design, a deep secondary settler must be provided to accommodate diurnal expansion of the biosolids blanket. Similar considerations apply to the design of the primary settler. Another point to consider is that systems not equipped with primary settling will possess greater nitrogen removal capacity.

In both aeration patterns, the last cell before the final settler should be an aerobic cell to nitrify any ammonia nitrogen released in the preceding anoxic cells and to purge any nitrogen gas clinging to the floc. If such an aerobic cell is not provided, settling difficulties may occur (32).

For more detailed discussion and information on stoichiometric and kinetic considerations of nitrification and denitrification, the reader is referred to Refs. (48–54).

4. MULTISTAGE SINGLE ANOXIC ZONE

4.1. Background and Process Description

The simplest continuous-flow single-sludge configurations rely on a dedicated compartment or tank for denitrification. The earliest investigation of single-sludge nitrification-denitrification processes for domestic wastewater was documented by Wuhrmann (42, 55), but a concurrent system was developed by Ludzack and Ettinger (46). These two systems are presented schematically in Figs. 7.3 and 7.4, respectively. The difference between these two systems is related to the carbon source for the denitrifying population. The Wuhrmann

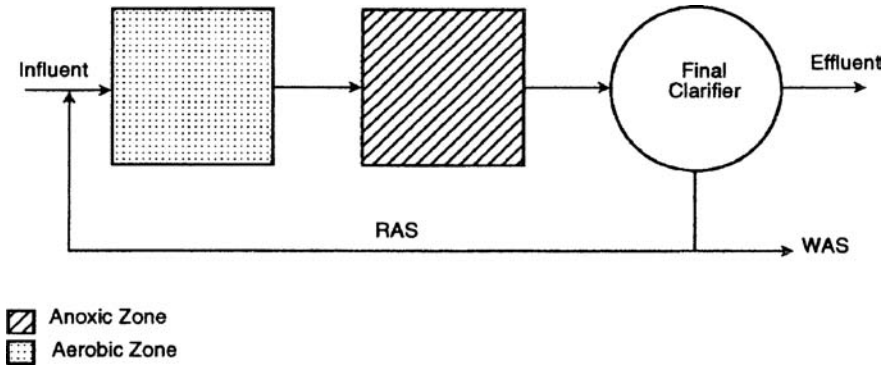

Fig. 7.3. Single anoxic zone-two stage Wuhrmann process. (*Source*: US EPA).

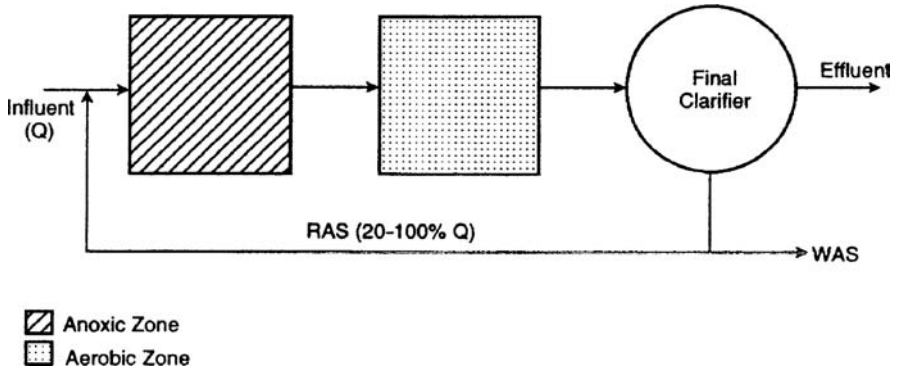

Fig. 7.4. Single anoxic zone-two stage Ludzack-Ettinger process. (*Source*: US EPA).

process places the denitrification reactor after the combined carbon oxidation/nitrification step, thus, this configuration has also been termed postdenitrification. The electron donor (carbon source) in a postdenitrification process train must be provided from endogenous decay, which is an intracellular depletion of organic carbon. The Wuhrmann process was not tested at full scale, but Christensen (56) was able to demonstrate 88% TN (total nitrogen) removal. Subsequent studies of the Wuhrmann process determined it to be unsuitable for full-scale application because of high effluent turbidities (presumably caused by lack of a post-aeration compartment, and/or long solids residence times (θ_c), the potential for increased effluent ammonia levels from lysed organisms, and low denitrification rates. The Wuhrmann design pioneered single-sludge nitrification–denitrification processes, but this process has not been used at full scale without modifications such as step-feed arrangements or supplemental carbon addition. Wuhrmann's effort provided the basic comprehension of the nitrification–denitrification process and microbiology for future refinements and modifications.

The system developed by Ludzack and Ettinger differed from the Wuhrmann system by placing the anoxic denitrification zone ahead of the aerobic zone, using external (exogenous) carbon provided by the raw wastewater. This type of process is termed predenitrification. The nitrate source was provided by directing the return activated sludge to the anoxic reactor. Conventional underflow ratios of 0.2:1 to 0.5:1 would not be expected to provide sufficient nitrates to optimize the amount of denitrification, and thus would be rate limiting.

Barnard (57) improved the Ludzack-Ettinger process by providing an additional internal MLSS (mixed liquor suspended solids) recycle from the aerobic stage to the anoxic stage to return nitrified MLSS at a regulated rate. This modification ensures adequate nitrates for the heterotrophic denitrification population. Process control and specific denitrification rates were enhanced with these modifications; consequently, process performance was improved. TN removals of 88% were achieved. The MLE (modified Ludzack-Ettinger), by Barnard, schematically illustrated in Fig. 7.5, was not extensively implemented at full scale, but was

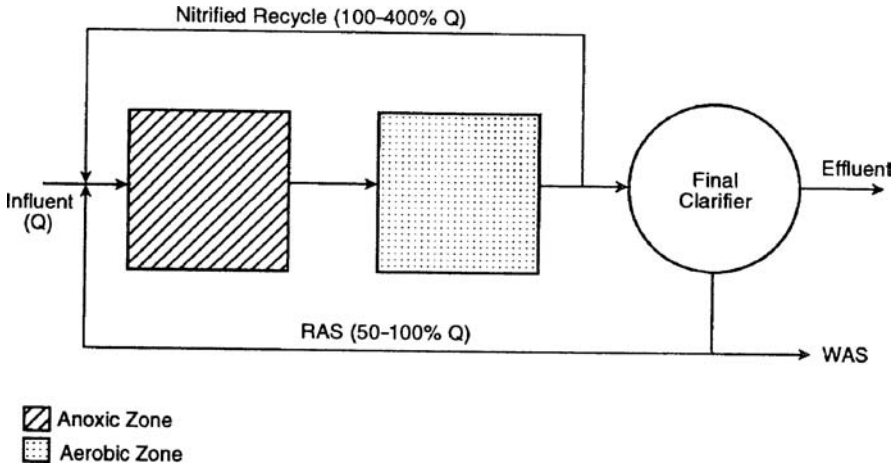

Fig. 7.5. Single anoxic zone modified Ludzack-Ettinger process. (*Source*: US EPA).

the progenitor of proprietary configurations, such as the A^2/O (Anaerobic/Anoxic/Oxic), Bardenpho, UCT (University of Capetown), and VIP (Virginia Initiative Plant). Variations of the MLE process design have been investigated by German and Japanese researchers (58–60). Schreiber and Menzel (59) proposed looped reactors, which place an anoxic reactor concentrically within the external aerobic reactor. Influent is received in the interior anoxic reactor, which may then be directed to the outer ring at the desired rate, either by a baffle system (Menzel process) or by a dedicated internal recycle line (Schreiber).

A proprietary single anoxic zone configuration is the A^2/O process, patented by Air Products, Inc. Originally developed for phosphorus removal as the A/O process (anaerobic/oxic), nitrification–denitrification was accommodated with the addition of an anoxic zone between the anaerobic and aerobic zones. Although the anaerobic zone is not required for nitrification–denitrification removal, it may be used at the start of the treatment train as an anaerobic "selector" for nitrification–denitrification in scenarios that do not require phosphorus removal. The anaerobic selector is used to control and maintain tank conditions to promote the profligation of zoogleal organisms, while suppressing the growth of filamentous organisms in the anoxic and aerobic reactors. Anoxic compartments located at the head end of the biological treatment train have demonstrated similar benefits (61). A schematic diagram of the A^2/O process is presented in Fig. 7.6.

The UCT process was developed at the University of Capetown in South Africa to surmount one of the inherent limitations of the MLE and A^2/O processes—the interference of nitrates on phosphorus removal processes. This was accomplished by: (1) returning activated sludge to the anoxic zone instead of to the anaerobic zone, and (2) providing an additional recycle from the anoxic zone to the anaerobic zone. The UCT process schematic is shown on Fig. 7.7. The purpose of these modifications is to denitrify nitrates returned by the RAS (return activated sludge) line before they are recycled to the anaerobic zone. A further refinement of the UCT process to accommodate lower strength wastewaters in the United States was investigated

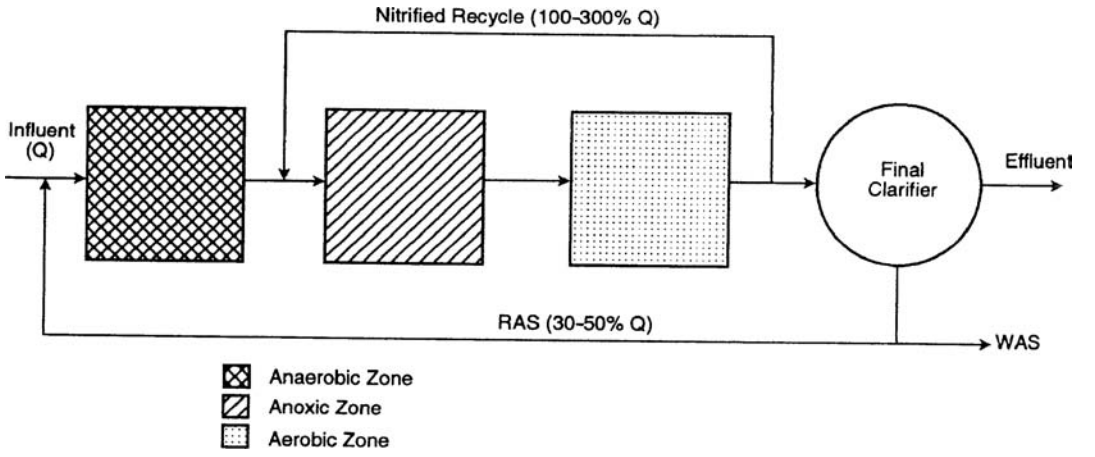

Fig. 7.6. Single anoxic zone A^2/O process with nitrification-denitrification. (*Source*: US EPA).

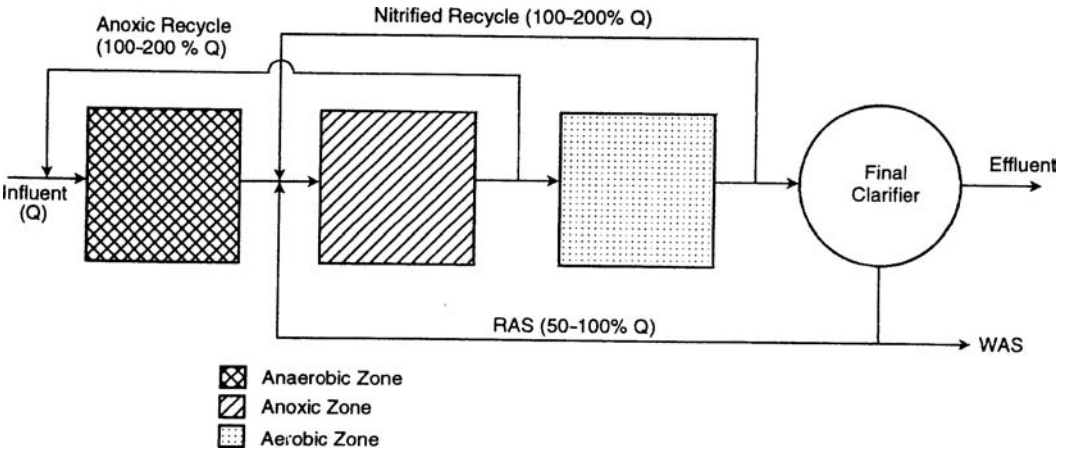

Fig. 7.7. Single anoxic zone University of Capetown (UCT) process. (*Source*: US EPA).

in Norfolk, Virginia. This process became known as the VIP process. Although the VIP and UCT processes are schematically similar, there are two fundamental differences: (1) the VIP process uses multiple complete mix cells instead of a single anaerobic reactor; this modification is intended to enhance phosphorus uptake by allowing a higher concentration of residual organics in the first anaerobic cell; and (2) because of the lower-strength wastewaters in the United States, a higher system rate (i.e., shorter θ_c) is afforded in the VIP process to increase the proportion of active biomass in the mixed liquor; this allows a smaller reactor volume and a shorter θ_c. The VIP process is patented, but its developers have waived the process fee.

4.2. Typical Design Criteria

Owing to their process limitations, the Wuhrmann and original Ludzack-Ettinger processes are not commonly used. The more recent predenitrification single anoxic zone processes are favored. Discussion of design criteria will be limited to the A^2/O, VIP/UCT, and MLE processes. Endogenous postdenitrification zones (e.g., those used in the Wuhrmann process) are used in some processes that employ multiple anoxic zones.

The design procedure for a single-sludge, single-anoxic zone nitrification–denitrification system consists of sizing the aerobic zone to nitrify the influent oxidizable TKN completely, and then sizing the anoxic zone and determining the required recycle rate (62–66).

The procedure for sizing the aerobic zone can be determined by conventional θ_c or nitrification rate considerations used in activated sludge nitrification applications, as discussed in Chapter 1. In summary, the sizing of the aerobic zone should consist of the following steps:

1. Select the design aerobic ${\theta_c}^d$.
2. Calculate secondary biosolids production.
3. Calculate the required aerobic zone solids inventory based on ${\theta_c}^d$.
4. Determine tank volume based on the solids inventory, settling properties, peaking factors, and the design MLSS.

Table 7.4
Design criteria for single anoxic zone predenitrification systems

Parameter	A^2/O	VIP/UCT	Generic single anoxic zone
MLSS, mg/L[a]	3000–5000	1500–3000	1500–4000
HRT, h			
Anaerobic[b]	0.5–1	1–2	0.5–2
Anoxic	0.5–1	1–2	0.5–2
Aerobic	3.5–6	2.5–4	2.5–6
θ_c, day	5–10	5–10	5–10
F/M, g BOD_5 applied/g MLVSS/day	0.15–0.25	0.1–0.2	0.1–0.3
RAS recycle, $\%Q$	20–50	50–100	50–100
Internal recycle			
Nitrified recycle, $\%Q$	100–200	200–400	100–400
Anoxic recycle, $\%Q$[c]		50–200	
Mix Power, hp/Mgal			
Anaerobic	50	70	40–70
Anoxic	50	70	40–70

Source: US EPA.

[a] Based on total mass of MLSS in all reactors. MLSS concentrations in individual compartments may vary because of the effect of recycle flows (RAS and IR) or step feeds.

[b] Only used in systems for both phosphorus and nitrogen removal or as a selector.

[c] A^2/O and MLE do not incorporate an anoxic recycle. Anoxic recycle is not required for systems that do not remove phosphorus.

The size of the anoxic zone should be based on the amount of nitrates to be denitrified. The required nitrate recycle rate is determined by the design effluent nitrate concentration.

The design criteria for the A^2/O process presented in Table 7.4 reflect data compilation from three full-scale plants. The design criteria presented for the VIP were obtained from the pilot-scale study performed at the Lamberts Point Wastewater Treatment Plant in Norfolk, Virginia, The VIP criteria differ from the A^2/O criteria because of the different objectives of each process and the conditions and influent characteristics at each site. The VIP process is designed to optimize nitrogen removal by providing two internal recycles. This modification affords a greater total recycle of nitrates for denitrification without affecting phosphorus removal processes. The A^2/O process is generally operated at a higher MLSS than the VIP and at a lower RAS rate. The lower RAS rates in the A^2/O process are required to ensure that the anaerobic selector is not overloaded with nitrates, which would adversely affect phosphorus removal.

4.3. Process Performance

Single anoxic zone systems will typically achieve total N effluent concentrations of <10 mg/L and long-term average effluent total N concentrations of 8 mg/L can reliably be achieved (9). Lower total N concentrations would require an additional anoxic zone or a separate denitrification step.

Higher recycles are required to achieve lower effluent nitrate concentrations. However, practical limitations on the recycle ratios, due to the energy required to pump large volumes detract from the viability of single anoxic zone technologies where effluent nitrogen limitations are $\leq$5 mg/L, or at facilities where >80% TN removal is required. Return biosolids rates are generally limited to 100% of the plant flow because of design solids considerations. Consequently, higher internal recycle rates are necessary to achieve lower effluent nitrogen levels. The increased capital and O&M costs and the effect of higher pumping rates on reactor retention time must be evaluated and compared with the benefit of enhanced nitrogen removal performance.

A plot of theoretical oxidizable nitrogen removal rate versus internal recycle for typical return biosolids rates (i.e., 50% to 100%) is illustrated in Fig. 7.8. The figure demonstrates that the maximum removal efficiency for a single anoxic reactor is 85% of oxidizable TN, at realistic recycle rates (i.e., $\leq$400%). These relationships do not consider denitrification that may occur in the final clarifier and within the floc particle in the aerobic zone, as hypothesized (55, 68). Consequently, observed nitrate removals may be greater than results predicted by theoretical considerations.

The A^2/O process has been implemented at the Largo Wastewater Treatment Plant in Florida. Performance data from that plant are presented in Table 7.5.

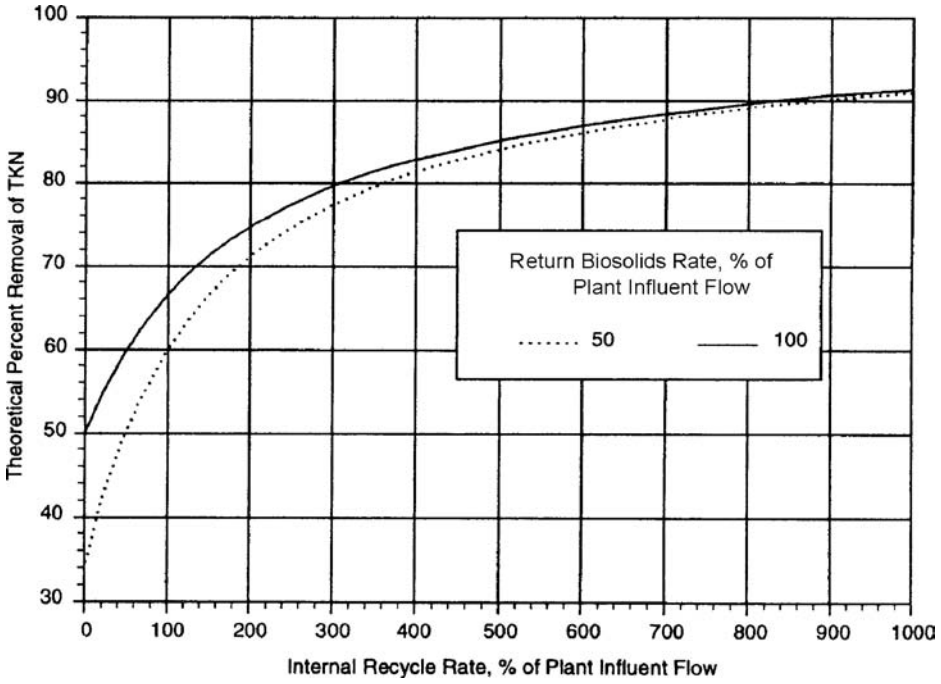

Fig. 7.8. Theoretical TKN removal for a single anoxic zone process as a function of internal recycle rate. (*Source*: US EPA).

Table 7.5
Performance summary of single anoxic zone processes

Parameter	A^2/O Largo, FL	VIP Pilot Norfolk, VA	MLE Landis, NJ
Q, m^3/day	39,360	151,400	19,300
BOD inf., mg/L	204	115	414
TKN inf., mg/L	23.5	24.4	34.7
BOD/TKN	8.7:1	4.7:1	11.9:1
TKN eff., mg/L	2.2	2.4	1.4
NH_4^+-N inf., mg/L	—	—	—
NH_4^+-N eff., mg/L	—	1.0	—
NO_3^--N eff., mg/L	5.7	5.3	4.4
Total N eff., mg/L	7.9	7.7	4.4
N removal, %	66	68	83

Source: US EPA.

The MLE process was also used at Maitland, Ontario, to treat a high-strength industrial wastewater (nitrate = 175 mg/L, NH_4^+-N = 190 mg/L, BOD_5 = 1230 mg/L). During optimum conditions, TN removals of 93% were obtained (68).

4.4. Process Design Features

Anoxic reactors most commonly use a continuously stirred tank reactor (CSTR) configuration; however, bench and pilot scale studies have investigated plug flow (69) and concentric circular reactors (58). These optional flow regimes did not appear to offer significant process improvement over the CSTR (59). Plug flow regimes offer better reaction kinetics; however, the increased oxygen demand for nitrification can result in organic overloading at the influent end of the reactor. This factor should be considered when designing the aeration system.

The division of a single aeration tank into anoxic and aerobic zones (and anaerobic zones for A^2/O) can be sufficiently achieved by a nonrigid baffle system.

Minimal DO should be introduced to the anoxic zone by influent and recycle flows or by surface transfer. Reduced denitrification rates at DO levels above 0.2 mg/L have been observed (70). Thus, nitrified internal recycle flow rates from the aerobic zone may require adjustment if excess DO is introduced in the anoxic zone. This problem can be mitigated in design by locating the internal recycle line inlet from the aerobic tank in a relatively unaerated corner of the tank where anoxic conditions may prevail. Also, submerged mixers should be designed not to entrain excessive air as a result of surface turbulence, but to provide sufficient mixing to ensure maximum dispersion and exposure of recycled nitrate and substrate to the denitrifying organisms.

Recycling of mixed liquor from the aerobic to the anoxic zone may typically involve high-volume, low-head pumping conditions. These applications may be achieved more economically by installing low-head submersible propeller pumps, wastewater pumps, or vertical turbine pumps directly in the aerobic basin, rather than by constructing a separate dry pit pump gallery.

Table 7.6
Monitoring requirements and rationale for single anoxic zone reactors

	Ludzack Ettinger, MLE		A^2/O, VIP, UCT	
Reactor	Parameter	Rationale	Parameter	Rationale
Anaerobic	N/A	N/A	DO, Nitrales	Presence of nitrates and DO will mitigate fermentive organisms
			Orthophosphates	Control to verify release
Anoxic	DO	Will reduce denitrification rate	DO	Will reduce denitrification rate
			NO_3	Inadequate load can cause excess phosphate release
	Q_1	Controls NO_3 load	Q_1	Controls NO_3 load
Aerobic	DO	High DO may inhibit denitrification; low DO may inhibit nitrification	DO	High DO may inhibit denitrification; low DO may inhibit nitrification
	Alkalinity, pH	Nitrification consumes alkalinity; may require pH control	Alkalinity, pH	Nitrification consumes alkalinity; may require pH control

Source: US EPA.

Both the MLE and A^2/O processes require only one MLSS internal recycle, thereby limiting process flexibility to only the RAS and internal recycle (IR). Additional flexibility and ability to bypass primary settling or step feed as needed may be achieved by providing interconnecting gates and channels. If phosphorus removal at the facility is required, RAS flow to the anaerobic zone must be minimized to limit nitrate interference. The UCT and VIP processes circumvent this limitation by conditioning RAS in the anoxic zone. This modification will enhance phosphorus removal and will also entail a higher degree of process monitoring, control, and operator sophistication. However, the nitrified recycle and RAS rates must be carefully controlled so that the nitrate load does not exceed the denitrification potential of the anoxic reactor, and result in a nitrate load to the anaerobic reactor that would cause a subsequent reduction of phosphorus removal.

A brief list of monitoring and control requirements for single anoxic zone systems is outlined in Table 7.6.

5. MULTISTAGE MULTIPLE ANOXIC ZONES

5.1. Background and Process Description

TN effluent concentrations $< 8\,mg/L$ cannot be consistently obtained using single anoxic zone processes without an additional attached growth filter or methanol supplement. TN effluent concentrations $< 6\,mg/L$ can be practically attained in a suspended growth system

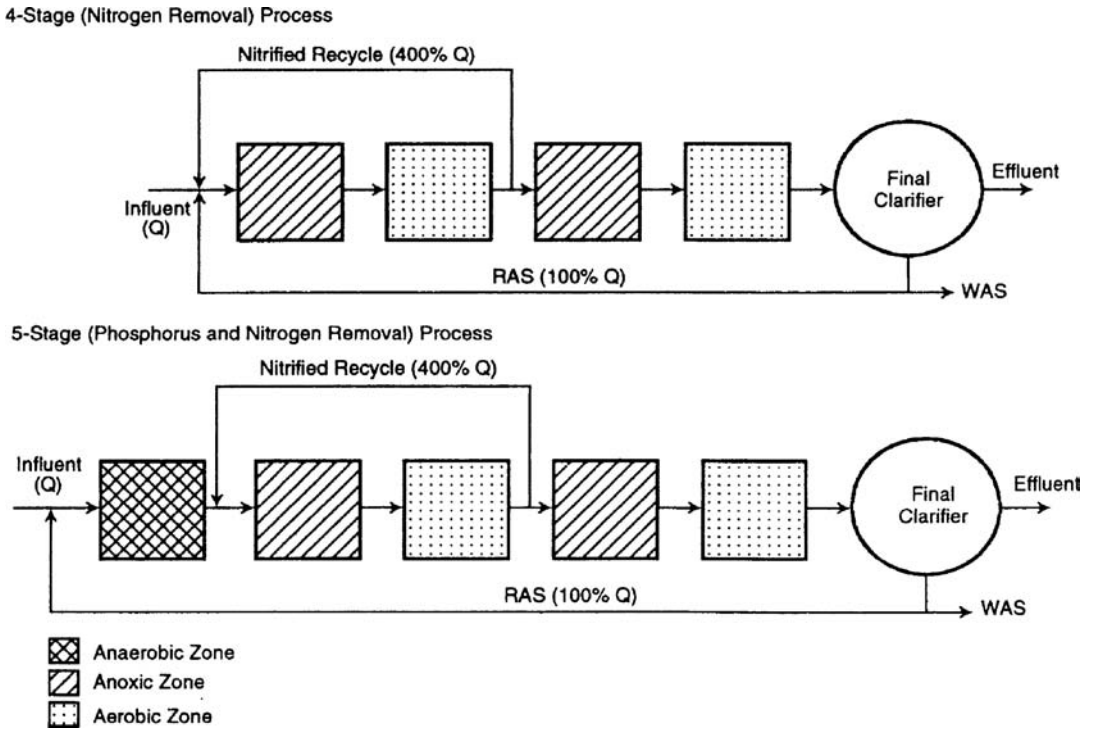

Fig. 7.9. Bardenpho process. (*Source*: US EPA).

without methanol addition by placing an endogenous anoxic zone in series after the aerobic zone. Although the A^2/O process does use two unaerated zones, the first (anaerobic) zone is not used for enhanced nitrogen removal but is provided for phosphorus removal or as an anaerobic selector. The first documented case of a second anoxic zone for denitrification was credited to Barnard, depicted schematically in Fig. 7.9. This process served as a precursor to the process he later patented as the Bardenpho process. Phosphorus removal was later accommodated in the Bardenpho process by placing an anaerobic reactor at the head of the treatment train, resulting in a five-stage process also illustrated in Fig. 7.9. The anaerobic fifth stage can be included in facilities that are not required to remove phosphorus as an anaerobic selector to suppress the growth of filamentous organisms.

The UCT process, described earlier, was also further modified by providing two anoxic zones (instead of one as in the original UCT) and two separate internal recycle lines. The purpose of this modification was to control the return biosolids (RAS shown in Fig. 7.10) and the nitrates recycle separately and also to reduce the nitrates load to the anaerobic reactor. Although the Modified UCT process uses dual anoxic zones, the second anoxic zone is not an endogenous denitrification reactor as was described for the Bardenpho process. Instead, the second anoxic zone in the Modified UCT is used only to denitrify recycled nitrates from the aerobic zone, and the first anoxic zone is exclusively used as an exogenous denitrification

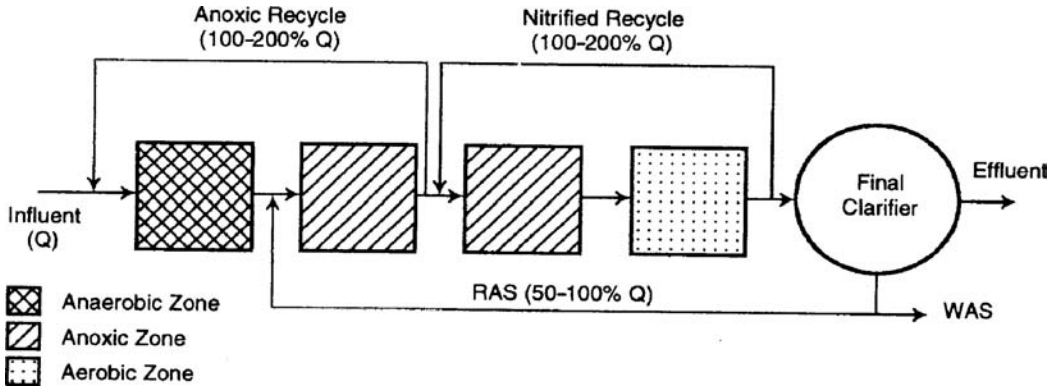

Fig. 7.10. Modified UCT process. (*Source*: US EPA).

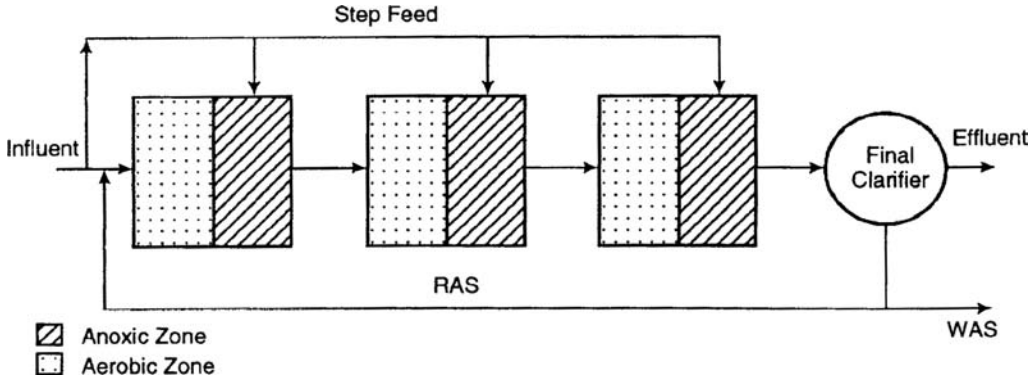

Fig. 7.11. Multi-anoxic zone with step feed. (*Source*: US EPA).

reactor to denitrify the RAS before recycle to the anaerobic zone. This allows increased recycle rates to the second anoxic zone for denitrification, and reduces nitrate interference of phosphorus removal in the anaerobic reactor.

A nonproprietary multianoxic zone process (60) is illustrated in Fig. 7.11. This design incorporated a three-stage sequence of aerobic-anoxic basins and a step feed to the second and third stages to supply the exogenous carbon source. The staging of the aerobic-anoxic zones served the purpose of an internal recycle, thereby offsetting O&M requirements with a larger capital cost associated with increased tank volume requirements. This configuration would presumably not offer the degree of process control compared to a design that included both IR and RAS.

The Bardenpho process is marketed in the United States by EIMCO. The patent describes a four-stage process, with one nitrified internal recycle and an activated sludge return. EIMCO administers a one time royalty fee for the process, which can include startup, training, and guarantee of performance (67).

5.2. Typical Design Criteria

Design criteria for single-sludge dual anoxic zone systems (i.e., MLSS, recycle rates, retention time, and mixing energy) are similar to criteria presented in a previous section for single anoxic zone systems. The most significant difference in dual anoxic zone design criteria from single anoxic zone design criteria relates to whether provisions for phosphorus removal are required. The long system θ_c's, which improve nitrogen removal, have been shown to adversely affect phosphorus removal. The four-stage Bardenpho, for instance, will typically be designed with a longer θ_c than configurations such as the A^2/O or VIP that are designed for phosphorus removal. The provision of a longer θ_c typically results in a lower biosolids production rate.

If phosphorus removal is desired, a five-stage Bardenpho can be selected by providing an anaerobic stage at the front of the four-stage Bardenpho treatment train. As a result, the first three stages of the five-stage Bardenpho process are similar to the A^2/O or VIP configuration. However, the final anoxic endogenous stage of the Bardenpho process affords two important process enhancements over processes that use single anoxic zones. The first is the additional degree of denitrification and consequent lower effluent TN concentrations. Second, the resulting reduced nitrate load to the final clarifier, which is recycled in the RAS to the anaerobic stage, reduces the potential for nitrate interference of phosphorus removal in the five-stage Bardenpho process. These features permit the use of higher internal recycle rates for a Bardenpho system than can be used with single anoxic zone systems that remove phosphorus and consequently improve nitrogen and phosphorus removal performance.

The procedure for sizing the first aerobic and anoxic zones of a dual anoxic zone process is identical to the procedures and concepts used for single anoxic zone systems. The first aerobic zone should be sized to nitrify the oxidizable influent TKN. The first anoxic zone of the Bardenpho system should be sized to completely denitrify the internal and RAS recycled nitrates. The first anoxic zone of a modified UCT process should be sized to denitrify nitrates in the RAS.

The second anoxic zone of a Bardenpho is sized to denitrify the nitrates not recycled to the first anoxic zone. The nitrate load to this zone is the difference between the oxidizable TKN and the nitrate reduced in the first anoxic zone. The tank volume will also be a function of the nitrate mass loading, temperature, MLSS, and SDNR (specific denitrification rate). Since endogenous denitrification rates are much slower than exogenous rates, the second basin will typically have a higher volume per mass of nitrates applied.

The second anoxic zone of a modified UCT process should be sized to denitrify the oxidizable TKN recycled from the aerobic reactor and the nitrates not recycled in the first anoxic zone.

Typical values used in the design of the four-stage Bardenpho and modified UCT systems are presented in Table 7.7.

The modified UCT has two fundamental process differences compared to a five-stage Bardenpho:

1. The modified UCT is designed to optimize phosphorus removal.

Table 7.7
Design criteria for dual anoxic zone

Parameter	4-Stage Bardenpho	Modified UCT
F/M, g BOD_5/g MLVSS/day	0.1–0.2	0.1–0.2
θ_c, day	10–40	10–30
MLSS, mg/L	2000–5000	2000–4000
HRT, h		
Anaerobic	—	1–2
1st Anoxic	2–5	2–4
Aerobic	4–12	4–12
2nd Anoxic	2–5	2–4
Reaeration	0.5–1	—
RAS, %	100	100
Internal recycle, %	400–600	100–600

Source: US EPA.

2. No endogenous denitrification is provided. Thus, the modified UCT would be unable to attain effluent TN concentrations consistently lower than 5 mg/L. Design for the modified UCT involves similar design concepts to a single anoxic zone process. Typically, the second anoxic zone of a modified UCT system is larger than the first anoxic zone due to the relationship of SDNR to influent COD (chemical oxygen demand). Since the COD to the second reactor is lower than to the first and less easily degradable, a lower SDNR will be experienced, necessitating a longer anoxic retention time. However, the rate in the second anoxic reactor will be greater than the endogenous rate in the second anoxic stage of the Bardenpho system.

The distinguishing characteristic of the modified UCT process is the complexity of internal recycling requirements, which exceed those of the Bardenpho design without offering a comparable degree of TN removal. As has been discussed, this lower degree of efficiency is caused by the phosphorus removal provision of the modified UCT.

To accomplish nitrification-denitrification without an internal recycle, primary effluent or raw wastewater can be step-fed to the anoxic zones. For a system such as that illustrated in Fig. 7.11, the optimum step-feed ratio can be derived or estimated from the influent wastewater characteristics. Each aerobic zone should be sized to completely nitrify all the influent TKN discharged to that zone. Likewise, the anoxic zone should be sized to completely denitrify the nitrates produced in the preceding aeration basin. The influent step feeds to the anoxic zones should be balanced such that the influent COD to each anoxic zone is sufficient to optimize exogenous nitrate respiration.

5.3. Process Performance

The Bardenpho design has achieved TN effluent concentrations of 3 mg/L and 90% removal afforded by the endogenous postdenitrification stage. The Bardenpho process has been used at several plants in the United States (71, 72). A list of typical performance data is included in Table 7.8.

In contrast, the modified UCT process has not been employed in the United States. Consequently, data for this configuration are unavailable for an assessment of the process.

Table 7.8
Summary of Bardenpho plant operating data

Plant	Flow, m^3/day (MGD)	Influent BOD_5, mg/L	Influent TKN, mg/L	Effluent Total N, mg/L	% N Removal
Tarpon Springs, FL	10,068 (2.66)	NA	NA	4.4	NA
Palmetto, FL	4656 (1.23)	160	36.60	2.9	92
Ft. Myers-Central, FL	23,429 (6.19)	135	23.30	2.7	88
Ft. Myers-South, FL	18,622 (4.92)	144	25.40	5.1	80
Payson, AZ	2574 (0.68)	196	32.80	3.2	90
Environmental Disposal Corp., NJ	818 (0.216)	190	17.20	2.8	84
Eastern Service Area, Orange County, FL	12,112 (3.2)	175	30.60	1.9	94
Kelowna, BC, Canada	12,491 (3.3)	188[a]	24.20	1.8	91
Hills Development, Pluckemin, NJ	908 (0.24)	169	18.3[b]	2.7	85

Source: US EPA.
[a] COD.
[b] NH_4^+–N only.

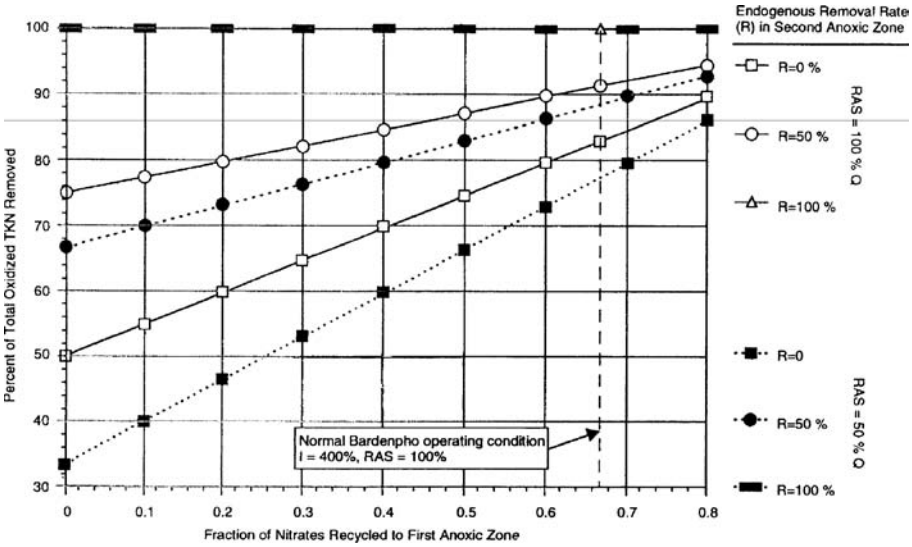

Fig. 7.12. Bardenpho process denitrification performance as a function of recycle rate. (*Source*: US EPA).

Percent removals as a function of the internal recycle rate are displayed in Fig. 7.12 for various endogenous removal rates and two RAS rates.

Figure 7.12 demonstrates that under normal conditions, the Bardenpho process can remove 83% of the oxidized TKN if no endogenous denitrification is considered. If 50% of the nitrates

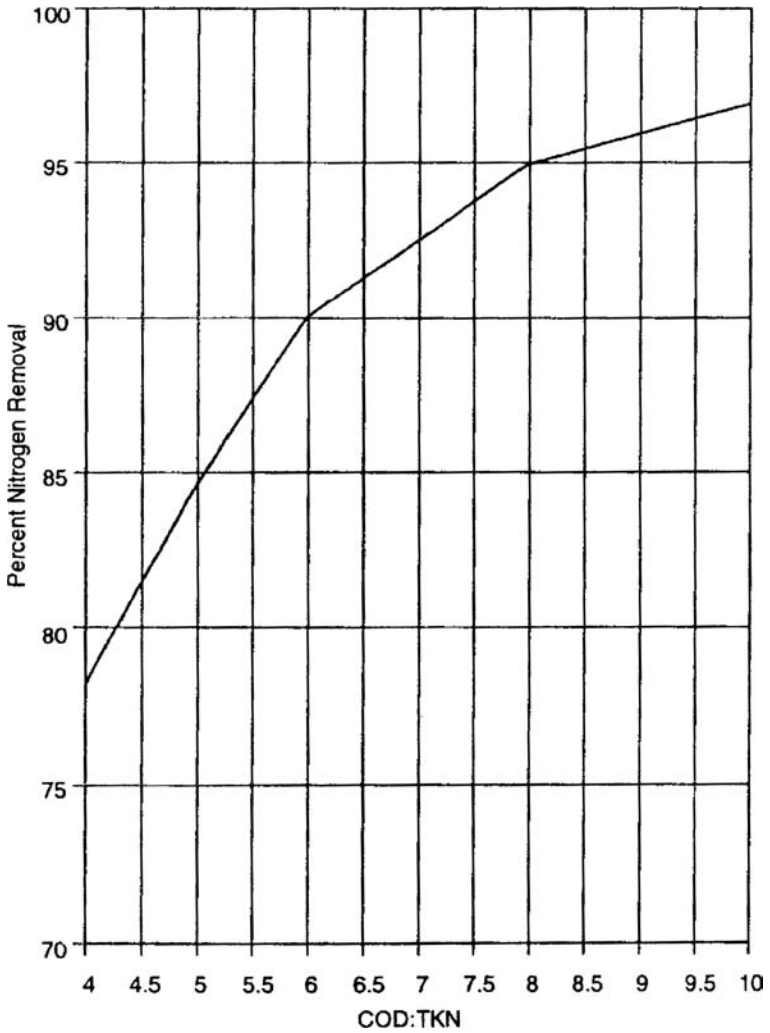

Fig. 7.13. Theoretical nitrogen removal as a function of COD:TKN for a triple anoxic zone process with step feed. (*Source*: US EPA).

to the second anoxic zone are removed through endogenous nitrate respiration (a conservative estimate), the nitrate removal performance increases to approximately 93%.

The theoretical removals as a function of the influent COD:TKN are presented in Fig. 7.13.

Figure 7.13 demonstrates step-feed processes that can theoretically achieve $>90\%$ removal if the COD:TKN $> 6:1$. The process can be further optimized by supplementing the last anoxic stage with methanol, or by providing a final endogenous reactor with postaeration.

5.4. Process Design Features

The Bardenpho process has been designed with plug flow, CSTR, and oxidation ditch flow regimes. However, the combined oxygen requirements of nitrification and carbonaceous oxidation can cause oxygen depletion in a plug flow aeration zone. The Bardenpho process incorporates many of the same process design features as the single anoxic zone processes. Design considerations should include the use of baffles for compartments, pump capacity for internal recycle requirements, and mixers in the anoxic zone to ensure maximum contact of nitrates and wastewater carbon with the micro-organisms. The Bardenpho process as a retrofit option was also determined to represent a viable option for existing plants that have their permits revised requiring nutrient removal (67). If sufficient tank volume exists, modifications may only require the installation of baffles and internal MLSS recycles. Plants that are not currently nitrifying may also require increased aeration capacity.

A five-stage Bardenpho plant should be designed to bypass the anaerobic zone in the event of a shock hydraulic or high DO load. For additional process operability and control, a prefermentation tank can be provided; alternatively, the anaerobic zone can be divided into compartments with baffles.

As was described for single anoxic zone systems, monitoring of the reactors in the Bardenpho process is required to ensure optimization of process performance. Suggested monitoring parameters and rationale are provided in Table 7.9.

6. MULTIPHASE CYCLYCAL AERATION

6.1. Background and Process Description

Alternating aerobic and anoxic zones can be achieved in a continuous-flow, activated sludge system by cycling the aerators on and off. This type of intermittent or pulsed aeration in an activated sludge facility is termed cyclical nitrogen removal (CNR). CNR processes can be most effectively applied at existing plants that have revised permits that impose nitrogen removal. Research and development of the CNR process has primarily been performed at a few existing plants, requiring only minor process modifications to convert to CNR. These modifications may be as minimal as installing baffles or timers to cycle aeration equipment, but may include providing internal recycle pumps and piping, or providing step-feeding capability. Thus, potential cost savings can be expected by implementing a CNR process when compared with conversion to a proprietary nitrogen removal process, if it is applicable.

TN removals of 80% in summer and just under 80% in winter were achieved at the Owego, New York, wastewater treatment plant (73). High θ_c, solids inventory control, and high COD:TKN were determined to be the key operational parameters. The process schematic for the Owego facility is presented in Fig. 7.14. Subsequent investigations (74) at the Barnstable, Massachusetts, wastewater treatment plant corroborated the Owego results.

An innovative alternating cyclical aeration process for nitrification-denitrification using countercurrent aeration is known as the Schreiber process (Fig. 7.15). The Schreiber process achieves alternating anoxic-aerobic zones within a single reactor by transferring air through submerged diffusers attached to a rotating arm. The mixed liquor typically rotates at a velocity

Table 7.9
Monitoring requirements and rationale for Bardenpho reactors

	Four-stage process		Five-stage (Phosphorus removal) process	
	Parameter	Rationale	Parameter	Rationale
Anaerobic	N/A	N/A	DO, Nitrates	Presence of electron acceptors will inhibit fermentive organisms
			Orthophosphates	Control to verify phosphate release
1st Anoxic	DO	Will reduce denitrification rate	DO	Will reduce denitrification rate
	NO_3	Inadequate load reduces amount of denitrification	NO_3	Inadequate load can cause excess phosphate release
	IR rate	Controls NO_3 load	IR rate	Controls NO_3 load
Aerobic	DO	High DO may inhibit denitrification rate; low DO may inhibit nitrification	DO	High DO may inhibit denitrification rate; low DO may inhibit nitrification
	Alkalinity, pH	Nitrification consumes alkalinity; may require pH control	Alkalinity, pH	Nitrification consumes alkalinity; may require pH control
2nd Anoxic	NO_3	High nitrification in aerobic zone may overwhelm endogenous denitrification capacity resulting in NO_3 in effluent	NO_3	High nitrification in aerobic zone may overwhelm endogenous denitrification capacity resulting in NO_3 in effluent
	DO	High DO will inhibit endogenous denitrification	DO	High DO will inhibit endogenous denitrification

Source: US EPA.

less than the moving bridge. The moving diffuser concept is intended to prevent bubble rise in a common vertical path and to prevent inducement of vertical currents. The manufacturer claims that this will maximize oxygen transfer by completely dispersing bubbles within the mixed liquor and increase the bubble detention time. Anoxic conditions can be achieved in the zone in front of the moving diffusers, while aerobic conditions exist in the zone immediately after the diffusers pass by that zone. Alternatively by using turbidity for process control, the single basin is cycled through oxic, anoxic, and anaerobic conditions. Mixing is maintained by the bridge rotation without aeration in the anoxic and anaerobic phases.

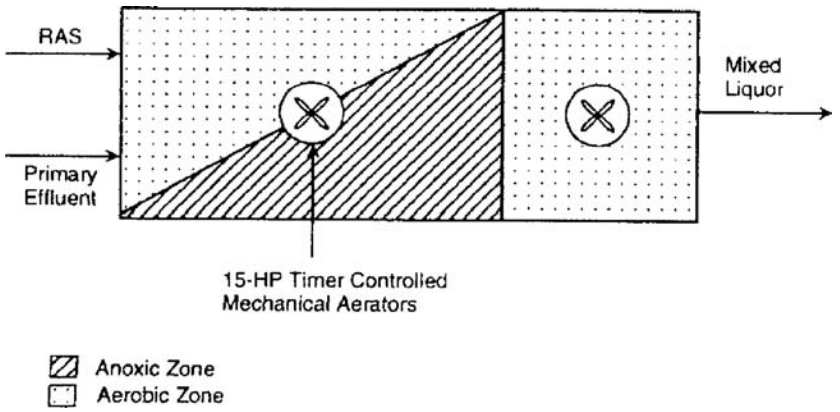

Fig. 7.14. Town of Owego, NY Water Pollution Control Plant. (*Source*: US EPA).

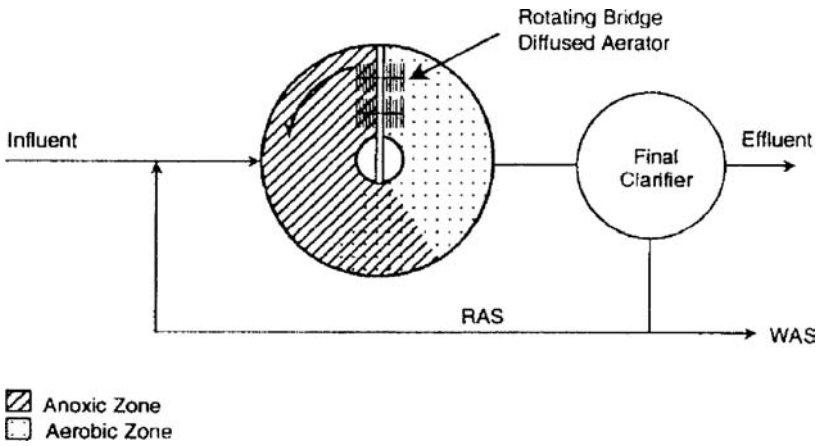

Fig. 7.15. Schreiber process. (*Source*: US EPA).

6.2. Typical Design Criteria

CNR design incorporates similar considerations as single anoxic zone processes. Aeration capacity, solids retention time (SRT or θ_c), solids inventory, and BOD:TKN are the most important design parameters. Bypassing primary settling to ensure a high COD:TKN for retrofit applications has been suggested (73); calculations should determine the adequacy of existing reactor basin volume, aeration capacity, and settling capacity. Design criteria are presented in Table 7.10.

CNR can be used to nitrify and denitrify without the use of an internal recycle. However, the capability to provide internal recycle should be considered as a process option or an on-demand basis.

Table 7.10
Cyclical aeration design criteria

Parameter	CNR at Owego	Schreiber
F/M, g/BOD_5/g MLVSS/day	0.06–0.13	0.05
Aerator on, min	15–45	*
Cycle off, min	15–30	*
θ_c, day	13–32	25
COD:TKN	10:1	
Aerobic DO, mg/L	1–1.5	0.5–1.5
Anoxic DO, mg/L	<0.3	
MLSS, mg/L	2600–4000	2000–7000

Source: US EPA.
* Load oriented with turbidity control.

Table 7.11
Cyclical aeration operating results

Process Location	CNR Barnstable, MA	CNR Owego, NY	CNR Blue Plains Wash., DC	Schreiber Clayton County, GA	Schreiber Jackson, TN
Q, m^3/day	5450	1820	N/A	8,970	31,260
HRT, h	9	13–16	10.1	N/A	N/A
θ_c, day	15	20–24	22.2	N/A	47.7
TKN in, mg/L	N/A	39.9	21.3	24.5	16.9
TKN out, mg/L	N/A	3.6	2.2	1.4	3.0
NH_4^+-N in, mg/L	22.3[a]	26.2	N/A	16	13.3
NH_4^+-N out, mg/L	3.2	1.4	1.0	0.5	1.2
N_{OX} out, mg/L	3.0	4.8	3.0	2.4	3.3
Total N removed, %	77[b]	80	76	84.5	63
COD:TKN	7.8[c]	10.5	9.3	N/A	
F/M, g BOD/g MLVSS/day	0.08[d] 0.24[e]	0.089	0.089		

Source: US EPA.
N/A = Data not available.
[a] Primary effluent.
[b] Based on influent NH_4^+-N only. Actual percent removed is higher, based on TKN.
[c] Ratio based on influent BOD to primary effluent TKN.
[d] Winter.
[e] Summer.

6.3. Process Performance

CNR systems can consistently produce effluent TN concentrations <8 mg/L and $>80\%$ TN removal. Pilot and full-scale operating results are presented in Table 7.11, along with full-scale operating results for the Schreiber process.

The CNR system offers flexibility, but requires more operator attention and expertise compared to other activated sludge modifications. Factors that introduce complexity to the process are the monitoring of nitrate, DO, and solids inventory, and adjustments in aeration cycles and step feeding that may be required to optimize nitrification and denitrification.

6.4. Process Design Features

Practical experience at full scale has suggested that the best performance for a continuous-flow, nonproprietary CNR system can be obtained using at least three basins in series and is recommended for design applications (73). The recommendation for a minimum of three basins in series is predicated on the provision of step feeding to the downstream basins. The process performance considerations indicated that process performance is enhanced by increasing the number of reactors.

Readers interested in modeling techniques and calibration of mathematical models for the simulation of nitrification and denitrification processes can find detailed information in Refs. (75–89).

7. PHOSPHORUS REMOVAL BY BIOLOGICAL AND PHYSICOCHEMICAL TECHNOLOGIES

7.1. Phosphate Biological Uptake at Acid pH

Various biotechnologies with or without chemical additions are discussed in this section. The readers are referred to the literature for detailed technical information (90–106) on phosphorus removal technologies.

Some micro-organisms store phosphate as polyphosphate, thereby removing it from solution, but current methods to induce this are unreliable. However, it was recently discovered that polyphosphate production is increased by acid shock. (90, 95). These studies have led to the identification of a significant, yet previously recognized, microbial stress response at acidic pH levels, which may be a novel strategy for the "one-step" removal of phosphate from wastewater effluents. It was possible to increase the level of phosphate removal by the microflora of a conventional activated sludge plant—under fully aerobic conditions—by more than 50% if the operational pH was adjusted within the range 5.5 to 6.5, as opposed to within 7.2 to 7.7, which are typical in current practice. Similar results were obtained in four other activated sludge plants of varying influent characteristics; enhancement of phosphate removal at pH 5.5 varied between 56% and 100% and involved a considerable fraction of the microbial population—bacterial, yeast, and fungal. Further research to assess the economic viability of a low-pH phosphate removal system is being carried out in Northern Ireland and Britain. (90).

7.2. Emerging Phosphorus Removal Technologies

The use of combined biological and physicochemical treatment processes for phosphorus removal was originally conceived by Beer and Wang at Coxsackie Sewage Treatment Plant, NY (51, 52, 83–87), and by Krofta and Wang at the Lenox Institute of Water Technology, MA (54, 101). They successfully used ferric chloride, lime, and alum for precipitation of phosphate from the activated sludge aeration basin effluent. Wang and Aulenbach (5) have discussed the

theory and principles of biological phosphorus uptake under aerobic conditions, biological phosphorus release under anoxic/anaerobic conditions, and physicochemical precipitation of released phosphorus (in phosphate form) by an innovative A/O process. Wang (102) has adopted a commercially available dissolved air flotation (DAF) clarifier in a combined biological physicochemical process system for high-rate phosphorus removal (104, 107).

Essentially, Wang's innovative technology (102) is a combined biological-chemical precipitation process involving the use of the following process steps:

1. Treatment of incoming wastewater (usually primary effluent) in an aeration basin to transfer phosphorus from the incoming wastewater to the "P-stripped activated sludge micro-organisms" under aerobic conditions.
2. Separation of the spent activated sludge micro-organisms from the aeration basin effluent by either a conventional sedimentation clarifier (with 2-hour definition time or DT), or by a high-rate dissolved air flotation (DAF) clarifier (with 15 minute DT) (54, 101, 102).
3. Discharge of the almost P-free clarifier effluent into a receiving water, while discharge of the P-concentrated clarifier sludge (either settled sedimentation clarifier sludge, or the floated DAF clarifier sludge) to a phosphate stripper (i.e., a thickener-type holding tank with 5 to 15 hour DT) where the clarifier sludge is subjected to anoxic/anaerobic conditions to induce phosphorus release from the clarifier sludge into the aqueous phase (102).
4. Chemical precipitation of the newly released phosphate in the phosphate stripper effluent (i.e., a P-rich, low-volume side stream containing 40 to 80 mg/L P; amounting to about 10% to 15% of the total wastewater flow) using lime, ferric chloride, or alum, and subsequently flotation of the P-rich precipitated chemical sludge for reuse as a fertilizer using a high-rate DAF clarifier (15 min DT).
5. Collection of the phosphate-stripped activated sludge, which has extremely high phosphorus uptake capacity from the stripper.
6. Return of the phosphate-stripped activated sludge to the aeration basin for reuse in a new cycle, where the phosphate-stripped activated sludge micro-organisms are again induced to take up dissolved phosphorus in excess of the amount required for growth under aerobic conditions (90).

The above P-removal process system reduces the volume of the wastewater to be treated (10% to 15% of total wastewater flow), thereby reducing the chemical dosage required, the amount of chemical sludge produced, and associated costs. Lime can be used to remove phosphorus from the stripper supernatant at lower pH levels (8.5 to 9.0) than normally required, although alum and ferric chloride are equally effective. The cycling of sludge through an anoxic phase may also assist in the control of hulking by the destruction of filamentous organisms to which hulking is generally attributed. The process is capable of reducing the total phosphorus concentration of typical municipal wastewaters to 1 mg/L or less. Adoption of a DAF clarifier instead of sedimentation clarifier for both secondary clarification and P-rich precipitated chemical sludge separation significantly reduces process time, and, in turn, saves overall capital and O&M costs. (102).

Wang (102) has also successfully used conventional biological sequencing batch reactors (SBR) instead of conventional activated sludge aeration for steps 1 and 6 above, and has used physicochemical sequencing batch reactors (PC-SBR) instead of the stripper and DAF for steps 3 and 4 above. The readers are encouraged to further improve upon this emerging phosphorus removal technology. Descriptions and discussions of PC-SBR can be found from the literature (17, 103, 107).

8. COXSACKIE WASTEWATER TREATMENT PLANT—A SINGLE-SLUDGE ACTIVATED SLUDGE PLANT FOR CARBONACEOUS OXIDATION, NITRIFICATION, DENITRIFICATION, AND PHOSPHORUS REMOVAL

8.1. Background Information

This project was initiated in 1970s when the New York State Department of Health (NYSDOH) and the New York State Department of Correction (NYSDOC) undertook a program of research and development in the area of advanced biological sewage treatment at one of its correctional facilities. With a view to the impending program of sewage treatment plant construction in the entire New York State, it was decided to build an advanced sewage treatment plant and a research laboratory on the grounds of the Coxsackie Correctional Facility at West Coxsackie, New York. The plant has been performing successfully to be a role model to the New York municipalities since 1973. This section only reports the detailed operational and and development data generated by Leo J. Hetling (PE, Ph.D.), Carl Beer (PE), and Lawrence K. Wang (Ph.D., PE), who were Research Director, Project Manager, and Chief Operator (NYSDEC Senior Sanitary Engineer), respectively, from 1973 to 1977. The plant data introduce the fact on how a typical single-sludge activated sludge plant performs for carbonaceous oxidation, nitrification, denitrification, and phosphors removal.

The Coxsackie Correctional Facility is an institution for young male delinquents aged 16 to 21. The average age of the inmates is l8. The design inmate population is 750. During the period covered by the full analytical data of this report, the inmate population was near capacity. In addition to the inmates, approximately 350 prison personnel are in daytime or nighttime residence at the facility. A 306 ha (750 acres) farming operation is part of the correctional facility. Farm products are milk, vegetables, apples, and beef. The institution is located in the south of Albany, NY. The effluent of the single-sludge biological treatment plant is discharged to the Coxsackie Creek, a short tributary of the Hudson River. The Coxsackie Creek is classified as an intermittent stream. The New York State effluent requirements for sewage treatment plant discharging to intermittent streams are as follows: (1) 5-day BOD = 5 mg/L, max., (2) ammonia nitrogen, NH_3 = 2 mg/L max., and (3) DO = 7.5 mg/L min.

The facilities that comprised the research installation are shown in Fig. 7.16. The main dimensions of the major treatment units employed on this project are summarized in Table 7.12.

8.2. Plant Operation and Parameters

The bulk of data describing the operating conditions and performance results are contained in Tables 7.12 to 7.23. The following remarks explain how some of these data were computed.

"Daily flow" is the daily flow introduced into the primary settler or the aeration tank via the variable sluice gate in the side weir structure, averaged over one calendar month. The daily flows for the month were added and the sum divided by the number of days in the month.

"Mean sewage detention time," for all units except the surge tank, is the average monthly nominal fluid retention time. It was found by dividing the "daily flow," as described above, into the volumetric capacity of the treatment unit. "Monthly high (low) sewage detention times"

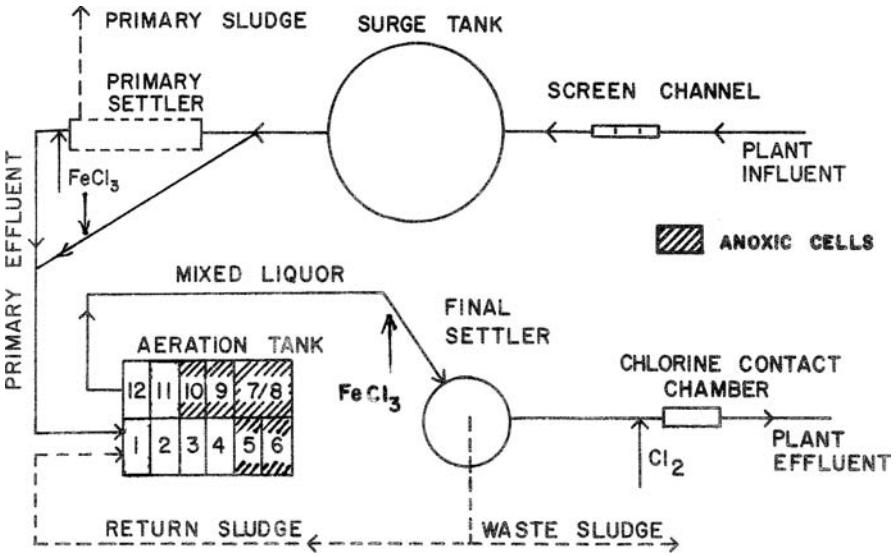

Fig. 7.16. Process schematic for Coxsackie sewage treatment plant, New York.

were determined by dividing the highest (lowest) daily plant flow of the calendar month into the capacity of the treatment unit.

Similar procedures were followed when determining the other hydraulic parameters based on plant flow: surface overflow rate and weir overflow rate.

"Mean sewage detention time" for the surge tank was determined as the flow weighted monthly average of the daily sewage detention time in the surge tank using the same method of computation used for determining the monthly flow weighted average of substrate concentrations found in the process water. "High" and "low" sewage detention times for the month were found simply by inspecting the list of daily detention times for the calendar month under consideration.

"Mixed liquor detention time" was computed by dividing the average monthly ML flow into the capacity of the aeration tank. ML flow is the sum of sewage flow and sludge recycle flow.

Monthly flow weighted average concentrations C_{m} were determined by using the following Eq. (12):

$$C_{\mathrm{m}} = \frac{\Sigma\, c_n Q_n}{\Sigma\, Q_n} \tag{12}$$

where

c_n = concentration, 24-hour composite, nth day of the month, mg/L
Q_n = daily flow, nth day of the month
Σ = summation for the days of the month

Table 7.12
Coxsackie Sewage Treatment Plant—Main dimensions of major treatment units and characteristics of equipment

AVERAGE DAILY FLOW:	
	570 m^3 (0.15 MGD)
SCREENS:	
	Two manually raked screens in series; 25 and 13 mm spacing of bars
GRIT REMOVAL:	
	Some grit removal occurs in surge tank and flow control structure (Side weir structure)
SURGE TANK:	
	9.76 m diameter; approximately 215 m^3 working capacity; equipped with floating aerator with direct-coupled, propeller-type impeller, 7.45 kw/1150 rpm; outpumping by submersible fixed-speed, torque-flow pumps to flow control structure
CONTROL OF FLOW TO TREATMENT UNITS:	
	Manual, by varying opening of submerged orifice in constant head device
PRIMARY SETTLER (Optional Use):	
	1.83 m wide; 7.32 m long; 2.44 m sidewater depth; 32.7 m^3 volume; 13.4 m^2 area; 1.83 m weir
AERATION TANK:	
	12 compartments, each 1.53 × 3.05 m in plan; sidewater depth 2.31 or 2.61 m, adjustable; volume under aeration 129 or 146 m^3; INKA aeration grids mounted near floor, 3.2 mm diam. airholes; one double compartment equipped with 0.25 kw/350 rpm mixer; compartments in series
AIR BLOWER:	
	Centrifugal type; 3.2 to 14.16 m^3/ min; 15 kw
FINAL SETTLING TANK:	
	Peripheral entry; 4.88 m diameter; 3.05 m sidewater depth; 57 m^3 volume; 18.70 m^2 area, 12.81 m weir
SLUDGE RETURN PUMPS:	
	Plunger Type
CHLORINE CONTACT CHAMBER:	
	15.9 m^3 volume

The same computational procedure was applied to temperatures and MLSS and RSSS concentrations.

The volatility of MLSS and RSSS in a single-sludge activated sludge system using iron and aluminum salts for phosphorus removal cannot readily be compared with the volatility of sludge occurring in a system not using this procedure. This was commented upon by many researchers (2–5, 49–54, 83–88). To permit better comparison of the volatilities achieved, the column "% Vol. Adjusted" was provided in Table 7.25. It was assumed that all iron in the sludge was present in the form of $Fe(OH)_3$ and that the molecular weight of $Fe(OH)_3$ is equal to 1.91 times the molecular weight of Fe (107/56 = 1.91).

Table 7.13
Flow-weighted, monthly average raw sewage characteristics for 12 months ending July 31, 1975

Parameter	Aug. 1974	Sept. 1974	Oct. 1976	Nov. 1974	Dec. 1974	Jan. 1975	Feb. 1975	Mar. 1975	Apr. 1975	May 1975	June 1975	July 1975
Daily Flow (m^3)	606	689	568	629	606	591	602	583	580	580	648	712
Daily Flow (million gal)	0.160	0.182	0.154	0.166	0.160	0.156	0.159	0.154	0.153	0.153	0.171	0.188
Temp. (°C)	N.AV.	26.5	26.7	24.9	21.7	23.6	21.0	26.5	24.3	24.1	26.1	28.1
pH, unit	N.AV.	7.2	7.5	7.7	7.7	7.4	7.5	7.3	7.3	7.3	7.3	7.2
Alk. (mg/L as $CaCO_3$)	N.AV.	161	169	182	207	185	196	160	174	178	177	158
SS (mg/L)	211	187	221	195	207	295	279	235	293	223	261	206
VSS (mg/L)	N.AV.	129	202	176	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.	N.AV.
COD (mg/L)	484	432	451	461	563	672	597	563	626	551	540	447
Diss. COD (mg/L)	209	186	233	231	N.AV.	293	243	264	287	293	323	229
BOD_5 (mg/L)	221	189	216	240	250	321	304	277	277	251	213	172
Diss. BOD_5 (mg/L)	111	94	122	112	N.AV.	155	145	148	140	138	123	83
NH_4^+-N (mg/L)	8.7	7.6	8.1	8.6	8.7	10.3	11.0	11.3	11.7	10.0	11.7	8.5
Org-N (mg/L)	18.9	14.2	14.3	15.6	18.3	20.7	18.0	17.9	18.2	18.5	18.8	13.9
TKN (mg/L)	27.5	21.9	22.1	24.2	26.9	31.0	28.5	29.9	29.9	30.1	30.5	22.4
NH_4^+-N as % of TKN	32	35	37	36	32	33	39	38	39	36	38	38
$NO_2^- + NO_3^- - N$ (mg/L)	1.0	1.3	0.7	0.7	0.8	0.4	0.4	0.4	0.5	0.5	0.3	0.4
Total N (mg/L)	28.5	23.2	22.9	24.9	27.7	31.5	29.5	30.2	30.3	30.6	31.4	22.9
TKN as % of COD	6	5	5	5	5	5	5	5	5	5	6	5
Ratio COD/TKN	18	20	20	19	21	22	21	19	21	18	18	20
Iron (mg/L)	0.3	0.3	1.5	0.3	0.6	0.4	0.4	0.2	1.1	1.6	1.2	1.8
Diss. Orthophosphorus (mg/L)	5.5	4.4	4.1	3.9	4.5	4.3	4.6	4.6	4.8	3.9	4.2	3.8
Diss. Phosphorus (mg/L)	7.7	6.2	4.9	4.5	5.8	5.6	6.0	6.3	6.9	5.4	5.5	4.3
Total Phosphorus (mg/L)	N.AV.	10.1	9.1	8.5	9.9	9.9	9.9	10.0	11.1	9.2	9.0	7.9
DO (mg/L)	N.AV.	5.3	4.8	5.4	5.6	5.3	5.8	5.0	5.0	5.8	5.0	4.0

Table 7.14
Monthly hydraulic surge tank, primary settler, and chlorine contact chamber data

	Surge Tank			Primary Settler				C.C.C.[b] S.D.T.[a]
	Sewage	Detention	Time	Surface Overflow Rate		Weir Overflow Rate		
Month Year	Mean (h)	High (h)	Low (h)	Mean ($m^3/m^2/day$)	Mean (gpd/ft^2)	Mean ($m^3/m/day$)	Mean (gpd/ft)	Mean (h)
Aug. 1974	8.2	10.9	5.5	N.AP.	N.AP.	N.AP.	N.AP.	0.7
Sept. 1974	8.2	11.7	5.6	N.AP.	N.AP.	N.AP.	N.AP.	0.6
Oct. 1974	8.4	12.1	3.3	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Nov. 1974	10.5	14.5	8.0	N.AP.	N.AP.	N.AP.	N.AP.	0.7
Dec. 1974	12.0	16.3	8.4	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Jan. 1975	11.2	21.4	6.6	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Feb. 1975	12.2	18.9	6.7	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Mar. 1975	11.9	20.7	4.2	N.AP.	N.AP.	N.AP.	N.AP.	0.8
Apr. 1975	11.1	15.8	6.1	35	(870)	260	(21,000)	0.8
May 1975	11.8	16.7	9.0	36	(880)	260	(21,000)	0.8
June 1975	11.1	16.8	6.9	43	(1050)	310	(25,000)	0.6
July 1975	8.6	12.8	5.9	47	(1160)	350	(28,000)	0.6

[a] S.D.T. = Sewage Detention Time.
[b] C.C.C. = Chlorine Contact Chamber.

Table 7.15
Aeration train profile data

Sampling Point		Sampling Time	DO (mg/L)	Temp. (°C)	pH	Total Alk. (mg/L as $CaCO_3$)	Dissolved TKN (mg/L)	NH_4^+-N (mg/L)	$NO_2^- + NO_3^-$-N (mg/L)	Dissolved Orthophosphorus (mg/L)
Surge Tank Effl.		8:39			7.5	173	17.08	12.0	0.2	2.2
Cell 1	Aerobic	9:06	3.2	23.5	7.2	135	12.32	9.4	2.4	0.77
Cell 2	Aerobic	9:33	3.9	23.4	7.2	136	10.08	8.5	4.0	0.42
Cell 3	Aerobic	10:00	3.4	23.2	7.1	117	7.28	5.5	5.0	0.25
Cell 4	Aerobic	10:27	4.1	23.0	7.0	118	5.60	4.1	7.1	0.24
Cell 5	Anoxic	10:54	0.1	23.0	7.1	122	5.60	4.3	4.8	0.21
Cell 6	Anoxic	11:21	0.1	23.0	7.1	129	5.60	4.6	2.7	0.20
Cell 7/8	Anoxic	12:13	0.1	22.5	7.1	143	6.44	5.1	0.1	0.49
Cell 9	Anoxic	12:39	0.1	22.5	7.3	142	5.88	5.0	0.1	0.25
Cell 10	Anoxic	1:05	0.1	22.5	7.2	144	5.88	5.0	0.1	1.08
Cell 11	Aerobic	1:31	5.8	22.1	7.3	124	3.64	2.4	2.4	0.33
Cell 12	Aerobic	1:57	6.5	22.2	7.2	109	1.96	0.4	0.1	0.15

Additional Surge Tank Effluent Data: Total P = 9.0 mg/L; Dissolved P = 2.3 mg/L; TKN = 30.2 mg/L. Additonal ML Data: MLSS = 3930 mg/L; MLVSS = 2600 mg/L; Flow-Through-Time = 26.5 min per cell (except Cell 7/8).
Fractional Rate of Return Sludge Recirculation = 0.44.

Table 7.16
Monthly average aeration tank and final settler detention times

		Aeration Tank						Final Settler		
		Sewage Detention Time			Mixed Liquor Detention Time			Sewage Detention Time		
Month	Year	Mean (hr)	Low (hr)	High (hr)	Mean (hr)	Low (hr)	High (hr)	Mean (hr)	Low (hr)	High (hr)
Aug.	1974	6.6	5.4	8.2	4.9	4.2	5.8	2.6	2.1	3.2
Sept.	1974	6.0	4.5	8.3	4.6	3.6	5.7	2.3	1.8	3.2
Oct.	1974	7.8	5.5	9.5	5.7	4.4	6.6	3.0	2.3	3.7
Nov.	1974	6.8	4.5	8.9	5.2	3.5	6.6	2.7	1.8	3.5
Dec.	1974	7.3	4.9	9.3	5.3	3.9	6.5	2.8	2.1	3.6
Jan.	1975	7.5	5.2	11.4	5.3	3.9	7.6	2.9	2.1	4.5
Feb.	1975	8.1	5.3	14.2	5.7	4.1	8.6	3.2	2.3	5.5
Mar.	1975	7.6	5.4	12.0	5.4	4.0	7.9	3.0	2.3	4.7
Apr.	1975	7.4	5.3	9.2	5.6	4.2	6.7	2.9	2.1	3.6
May	1975	7.3	5.2	9.6	5.5	3.9	7.1	2.9	2.0	3.7
June	1975	6.1	4.8	9.5	4.7	3.8	6.8	2.4	1.9	3.7
July	1975	5.5	4.8	7.4	4.3	3.7	5.6	2.1	1.9	2.9

Table 7.17
Monthly average final settler overflow and solids loading rates

		Surface Overflow Rate ($m^3/m^2/day$)			Surface Overflow Rate (gpd/ft^2)			Weir Overflow Rate		Solids Loading Rate[a]	
								($m^3/m/day$)	(gpd/ft)	($kg/m^2/day$)	($lb/ft^2/day$)
Month	Year	Mean	High	Low	Mean	High	Low	Mean	Mean	Mean	Mean
Aug.	1974	29	35	23	(711)	(850)	(565)	41	(3300)	113	(23)
Sept.	1974	31	42	23	(760)	(1030)	(555)	46	(3700)	113	(23)
Oct.	1974	24	34	20	(588)	(835)	(485)	35	(2800)	118	(24)
Nov.	1974	27	42	21	(662)	(1020)	(520)	40	(3200)	100	(20)
Dec.	1974	26	38	20	(638)	(940)	(495)	37	(3000)	117	(24)
Jan.	1975	25	36	17	(613)	(890)	(405)	36	(2900)	129	(26)
Feb.	1975	23	35	13	(564)	(865)	(325)	34	(2700)	105	(21)
Mar.	1975	25	35	16	(613)	(860)	(385)	36	(2900)	125	(26)
Apr.	1975	25	35	20	(613)	(865)	(500)	37	(3000)	120	(25)
May	1975	26	36	20	(638)	(890)	(480)	37	(3000)	125	(26)
June	1975	31	39	20	(760)	(960)	(485)	45	(3600)	143	(29)
July	1975	34	39	25	(834)	(965)	(620)	49	(3900)	167	(34)

[a] Solids Load Rate $= \frac{\text{MLSS (mg/L)} \times \text{Mixed Liquor Flow } (m^3/day)}{\text{Final Settler Area } (m^2) \times 10^3}$; $kg/m^2/day = \frac{8.34 \text{ MLSS (mg/L)} \times \text{Mixed Liquor Flow (MGD)}}{\text{Final Settler Area } (ft^2)}$; $lb/ft^2/day$.

Table 7.18
Flow-weighted, monthly average surge tank effluent characteristics

Parameter	Aug. 1974	Sept. 1974	Oct. 1974	Nov. 1974	Dec. 1974	Jan. 1975	Feb. 1975	Mar. 1975	Apr. 1975	May 1975	June 1975	July 1975
Daily Flow (m^3)	534	583	451	511	477	466	432	458	473	477	572	633
Daily Flow (mil gal)	(0.141)	(0.154)	(0.119)	(0.135)	(0.126)	(0.123)	(0.114)	(0.121)	(0.125)	(0.126)	(0.151)	(0.167)
Temperature (°C)	26.0	24.1	23.8	20.7	18.0	18.8	17.8	20.5	19.7	23.1	24.8	27.6
pH, unit	7.2	7.2	7.2	7.4	7.4	7.4	7.5	7.4	7.4	7.4	7.5	7.4
Alk. (mg/L as $CaCO_3$)	168	179	190	204	219	212	229	206	197	212	217	172
SS (mg/L)	205	241	270	223	276	241	242	267	250	246	253	218
COD (mg/L)	413	378	433	398	396	434	457	464	457	434	429	386
BOD_5 (mg/L)	181	163	195	186	194	208	225	213	189	177	157	142
Diss. BOD_5 (mg/L)	63	59	66	53	N.AV.	92	75	75	73	60	47	40
NH_4^+-N (mg/L)	10.5	9.6	10.9	12.3	12.0	12.8	14.9	14.7	13.7	16.3	15.1	11.0
Organically Bound N (mg/L)	15.6	12.3	14.8	12.1	14.5	15.6	17.1	15.5	15.4	13.9	13.9	12.3
TKN (mg/L)	26.0	22.3	26.1	24.4	26.2	28.4	32.0	30.2	29.3	30.3	29.0	23.3
NH_4^+-N as % of TKN	41	43	42	50	46	45	47	49	47	54	52	47
$NO_2^- + NO_3^-$-N (mg/L)	0.25	0.2	0.01	0.05	0.14	0.07	0.13	0.13	0.25	0.31	0.24	0.15
Total N (mg/L)	26.0	22.5	26.1	24.5	26.3	28.7	32.1	30.4	29.5	30.6	29.2	23.5
TKN as % of COD	6	6	6	6	7	7	7	7	6	7	7	6
Ratio COD/TKN	16	17	17	16	15	15	14	15	16	14	15	17
Diss. Orthophosphorus (mg/L)	6.4	4.3	4.0	4.5	4.9	4.6	5.3	5.5	5.5	4.5	4.4	4.0
Diss. Phosphorus (mg/L)	7.1	4.6	4.3	4.4	4.9	4.6	5.0	5.5	4.1	N.AV.	N.AV.	N.AV.
Total Phosphorus (mg/L)	11.4	9.7	9.9	9.1	10.3	9.6	10.2	10.8	10.0	9.2	9.1	8.3
DO (mg/L)	N.AV.	0.3	0.4	0.4	0.4	0.3	0.4	0.4	0.3	0.3	0.2	0.3

Table 7.19
Monthly average COD and BOD5 removal efficiencies

		COD Removal					BOD_5 Removal				
Month	Year	Surge Tank (%)	Primary Settler (%)	Aeration System (%)	Chl. Cont. Chamber (%)	Total Plant (%)	Surge Tank (%)	Primary Settler (%)	Aeration System (%)	Chl. Cont. Chamber (%)	Total Plant (%)
Aug.	1974	14	N.AP.	92	N.AP.	N.AP.	17	N.AP.	97	N.AP.	N.AP.
Sept.	1974	10	N.AP.	88	30	92	14	N.AP.	95	62	98
Oct.	1974	4	N.AP.	91	13	93	10	N.AP.	97	40	99
Nov.	1974	14	N.AP.	87	15	90	22	N.AP.	95	56	98
Dec.	1974	30	N.AP.	74	5	83	22	N.AP.	84	13	90
Jan.	1975	35	N.AP.	89	4	93	35	N.AP.	93	36	97
Feb.	1975	23	N.AP.	87	8	91	26	N.AP.	91	43	96
Mar.	1975	16	N.AP.	87	13	90	22	N.AP.	91	44	96
Apr.	1975	27	22	81	8	90	32	18	92	58	98
May	1975	18	26	81	10	90	27	28	91	36	97
June	1975	21	31	82	7	91	26	30	93	75	99
July	1975	14	36	79	10	90	17	38	98	0	99

For treatment units, efficiency is based on unit influent contaminant concentrations; for entire plant, efficiency is based on raw sewage contaminant concentrations.

Table 7.20
Effect of ferric chloride addition on process phosphorus removal and iron concentration of return sludge

Month	Year	Total P in Influent to Aeration Train (mg/L)	Fe^{3+} Dosage (mg/L)	Molar Ratio Fe/P	Fe^{3+} Content of Return Sludge (%)	Total P Removed (%)
Aug.	1974	11.5	12.5	0.60	8.6	83
Sept.	1974	9.7	13.7	0.78	8.4	92
Oct.	1974	9.9	8.7	0.49	5.4	95
Nov.	1974	9.1	7.8	0.47	4.3	82
Dec.	1974	10.3	8.0	0.43	4.1	55
Jan.	1975	9.6	12.6	0.73	5.8	79
Feb.	1975	10.2	9.0	0.48	3.7	69
Mar.	1975	10.6	15.1	0.79	6.7	81

The following equation was used for computing the values in the column headed "% Vol. Adjusted":

$$\%\ \text{vol. adjusted} = \left(\frac{\%\text{vol.}}{100 - \%\,\text{Fe} \times 1.91}\right) 100 \tag{13}$$

where:

% vol. = percent volatile solids determined
% Fe = percent Fe^{3+} found in (total) MLSS or RSSS.

In other words, the volatile solids mass is related to the total solids mass minus the mass of the assumed iron compound therein.

Sewage detention times in the surge tank are presented in Table 7.15. Dissolved BOD decreased by 60 mg/L; particulate BOD increased by 5 mg/L.

This flow sheet, shown diagrammatically in Figs 7.2 and 7.16 and characterized by the omission of primary settling and the use of a single anoxic zone in the aeration train comprising Cells 5, 6, 7/8, 9, and 10, was evaluated by Hetling, Beer, and Wang (51–54, 83–88). The aerator in the surge tank was on a 2 mm "on," 18 mm "off" cycle. Several modes of ferric chloride dosing were employed. Operating conditions and performance results for the Flow Sheet shown in Fig. 7.16 are reflected in Tables 7.14 to to 7.25. Provided that a system has sufficient nitrogen removal capacity via sludge synthesis and denitrification, nitrogen removal efficiency will depend on the flocculating properties of the sludge and will parallel SS and BOD_5 removals. Poor flocculating properties will be reflected in high SS in the effluent, paralleled by high BOD_5 and TKN concentrations.

Table 7.21
Flow-weighted, monthly average chlorine contact chamber effluent characteristics[b]

Parameter		Aug. 1974	Sept. 1974	Oct. 1974	Nov. 1974	Dec. 1974[a]	Jan. 1975	Feb. 1974	Mar. 1975
Daily Flow, Surge Tank Effluent	(m^3)	534	583	451	511	477[a]	466	432	458
Daily Flow, Surge Tank Effluent	(million gallons)	(0.141)	(0.154)	(0.119)	(0.135)	(0.126)[a]	(0.123)	(0.114)	(0.121)
Temperature of Chl. Cont. Chamber Effluent	(°C)	N.AV.	22.2	20.7	18.9	15.6[a]	15.8	15.0	17.2
Temperature of Air	(°C)	20.5	14.3	5.8	3.9	−1.1	−2.8	−2.2	−0.6
DO	(mg/L)	N.AV.	3.8	4.5	4.9	5.2[a]	4.9	4.3	4.9
pH	(unit)	7.0	7.1	7.0	7.1	7.2[a]	7.0	7.2	7.1
SS	(mg/L)	4.0	6	10	18	60[a]	16	16	11
COD	(mg/L)	N.AV.	33	33	44	96[a]	46	56	53
Diss. COD	(mg/L)	N.AV.	28	28	29	N.AV.	37	36	47
% Diss. COD	(%)	N.AV.	85	85	66	N.AV.	80	64	88
BOD_5[c]	(mg/L)	N.AV.	3	3	4	26[a]	9	12	9
Diss. BOD_5[c]	(mg/L)	N.AV.	3	2	2	N.AV.	5	4	7
% Diss. BOD_5	(%)	N.AV.	100	67	50	N.AV.	56	33	78
NH_4^+-N	(mg/L)	0.3	0.5	0.1	1.6	3.1[a]	0.7	1.1	2.4
$NO_2^- + NO_3^-$-N	(mg/L)	1.5	1.3	1.2	1.2	2.2[a]	1.2	1.0	1.6
Diss. Orthophosphorus	(mg/L)	1.9	0.4	0.3	0.7	2.1[a]	1.0	2.1	1.2

[a] December data are of little significance because plant was operated as a primary plant for part of the month due to breakdown of final settler equipment.
[b] Average chlorine dose was 6.2 mg/L.
[c] BOD_5 samples were seeded before analysis.

Table 7.22
Stoichiometric data abstracted from nitrogen profiles

	Date of Profile						
Parameter	11/14/73	11/27/73	12/5/73	9/18/74	10/10/74	11/14/74	6/12/75
Temperature of process water (°C)	20	20	18	23	22	20	25
Nitrification (mg/L of NH_4^+-N oxidized)	10.8	10.8	11.2	6.9	8.7	5.8	18.9
Alkalinity destruction during nitrification (mg/L)	N.AV.	60.0	82.0	N.AV.	52	45	130
Specific alkalinity destruction (mg/mg)	N.AV.	5.6	7.3	N.AV.	6.0	7.8	6.9
Denitrification via ENR (mg/L N gasified)	7.8	7.8	8.0	4.6	7.0	3.7	12.2
Alkalinity release during denitrification (mg/L)	N.AV.	20	28	N.AV.	26	8	48
Specific alkalinity release (mg/mg)	N.AV.	2.6	3.5	N.AV.	3.7	2.2	3.9
Ammonia release (Uptake) during ENR (mg/L)	(0.2)	0.2	0.2	0.8	0.9	(0.3)	2.1
Specific ammonia release during ENR	Q.R.	0.026	0.025	0.174	0.129	Q.R.	0.173

Table 7.12 introduces the main dimensions of major treatment units and characteristics of equipment at Coxsackie plant, shown in Figs. 7.2 and 7.16. Table 7.13 presents the flow-weighted, monthly average raw sewage characteristics for one full year (August 1974 to July 31, 75). Figure 7.17 shows a typical flow pattern of raw sewage at the Coxsackie Sewage Treatment Plant. The raw sewage is treated by a surge tank and a primary settler before it reaches the aeration. Table 7.14 shows hydraulic conditions of the surge tank, primary settler, and chlorine contact chamber, whereas Table 7.15 shows the aeration train profile data for the flow sheet (Fig. 7.16). The readers are referred to the dissolved oxygen (DO) concentrations maintained in the aeration compartments (Fig. 7.2 and Table 7.15). The aeration effluent flows to a final settler and a chlorine contact chamber (Fig. 7.16). Table 7.16 indicates the monthly average aeration tank and final settler detention times. Table 7.17 indicates the monthly average final settler overflow and solids loading rates.

Table 7.23
Kinetic data abstracted from nitrogen profiles[b]

		Date of Profile						
		11/14/73	11/27/73	12/5/73	9/18/74	10/10/74	11/14/74	6/12/75
Temperature of Process Water[a] (°C)		20	20	18	23	22	20	25
Nitrification Rate[c] (mg NH_4^+-N oxidized/L reactor volume/h)	Highest	5.0	7.2	7.2	4.6	4.8	3.6	15.9
	Average	3.56	4.14	4.34	2.71	4.02	2.28	11.00
Nitrification Rate[c] (kg NH_4^+-N oxidized/kg MLVSS/day)	Highest	0.055	0.079	0.070	0.062	0.044	0.055	0.146
	Average	0.039	0.045	0.042	0.036	0.037	0.035	0.102
Denitrification Rate (mg NO_3^--N reduced/L reactor volume/hr)	Highest	3.9	4.5	4.6	3.0	5.2	4.6	4.9
	Average	3.23	3.23	3.31	2.65	3.96	2.22	3.73
Denitrification Rate (mg NO_3^--N reduced/g MLSS/h)	Highest	1.19	1.24	1.14	1.07	1.33	1.71	1.19
	Average	1.00	0.90	0.83	0.95	1.01	0.83	0.90
Denitrification Rate (kg NO_3^--N reduced/kg MLVSS/day)	Highest	0.043	0.049	0.044	0.040	0.048	0.069	0.046
	Average	0.035	0.035	0.032	0.035	0.037	0.034	0.034

[a] Temperatures are in part estimated in accordance with monthly average temperatures.
[b] When computing average rates of denitrification only four cells were counted as operative anoxic cells; the remaining two anoxic cells were inoperative due to exhaustion of nitrates.
[c] Average rate of nitrification refers to the aerobic cells placed upstream of anoxic cells.

8.3. Performance Results

8.3.1. Carbonaceous Oxidation

Tables 7.12 to 7.17 and Fig. 7.17 document the operational conditions of advanced Coxsackie plug flow single sludge treatment plant. Tables 7.18 to 7.26 and Figure 7.18 and 7.19 document the plant's performance data in one year period for carbonaceous oxidation, nitrification, denitrification, phosphorus removal, and sludge chlorination treatment..

As shown in Tables 7.19 and 7.21, Coxsackie plant performs satisfactorily for carbonaceous oxidation. COD reduction and BOD_5 reductions are in the range of 81% to 92% and 90% to 99%, respectively.

Table 7.24
Monthly rates of denitrification abstracted from monthly nitrogen balances

Month	Temp. Final Settler Effl. (C)	ML Vol. Fract. (%)	Rates of Denitrification[a] (mg/g VSS/h)	(g/g VSS/day)	(mg/g VSS/h)[b]	(g/g VSS/day)[b]	Growth Rate-HM (day^{-1})
Aug.'74	25.5	65	2.09	0.050	3.14	0.075	0.23
Sept.'74	22.5	63	1.60	0.038	2.40	0.058	0.28
Oct.'74	21.5	65	1.22	0.029	1.83	0.044	0.18
Nov.'74	19.3	69	1.07	0.026	1.61	0.039	0.26
Dec.'74[c]	16.3[c]	69[c]	0.89[c]	0.021[c]	1.34[c]	0.032[c]	0.21[c]
Jan.'75	16.6	71	1.09	0.026	1.64	0.039	0.16
Feb.'75	15.9	73	1.24	0.030	1.86	0.045	0.22
Mar.'75	18.1	68	1.28	0.031	1.92	0.046	0.19
Apr.'75	18.5	62	1.30	0.031	N.AP.	N.AP.	0.16
May'75	23.0	65	1.51	0.036	N.AP.	N.AP.	0.12
June'75	24.5	62	1.82	0.044	N.AP.	N.AP.	0.14
July'75	27.1	59	1.71	0.041	N.AP.	N.AP.	0.11

[a] Rates of denitrification refer to N removed in ENR.
[b] Adjusted for two ineffective anoxic cells in the aeration train, HM . . . heterotrophic matrix.
[c] December data are of little significance because plant was operated as primary plant for part of the month due to breakdown of final settler equipment.

Table 7.25
Monthly average excess activated sludge characteristics

	RSSS (mg/L)	RSVSS (mg/L)	Fe^{3+} (mg/L)	Fe^{3+} (% of RSSS)	TKN (mg/L)	TKN (% of RSSS)	P[a] (mg/L)	P (% of RSSS)	Molar Ratio Fe/P	RSSS/ MLSS	% Vol.	%[c] Vol. Adj.
1974												
Aug.	11,400	7200	940	8.6	880	5.3	559	3.9	0.93	3.84	63	75
Sept.	11,100	6900	930	8.4	640	5.8	524	4.7	0.98	3.96	62	74
Oct.	13,300	8500	710	5.4	770	5.8	671	5.1	0.59	3.71	64	71
Nov.	12,000	8100	520	4.3	710	5.9	469	3.9	0.62	4.29	68	74
Dec.	13,300	8500	550	4.1	750	5.6	511	3.8	0.59	4.00	64	69
1975												
Jan.	13,100	9200	760	5.8	840	6.4	475	3.6	0.88	3.56	70	79
Feb.	11,300	8300	420	3.7	750	6.7	368	3.3	0.64	3.52	73	79
Mar.	12,700	8600	810	6.7	680	5.7	537	4.5	0.81	3.52	67	77
Apr.	15,000	9500	1320	8.8	810	5.4	620	4.1	1.18	4.21	63	76
May	14,200	11,400	1050	7.4	850	6.0	600	4.2	0.95	3.85	65[b]	76
June	14,500	9100	1290	8.9	780	5.4	713	4.9	1.00	4.00	63	76
July	16,600	10,300	1690	10.1	890	5.4	833	5.0	1.12	4.31	62	77

[a] Determined by computation.
[b] Determined from ML data.
[c] Before computing % Vol., weight of Fe^{3+} as $Fe(OH)_3$ was deducted from RSSS.

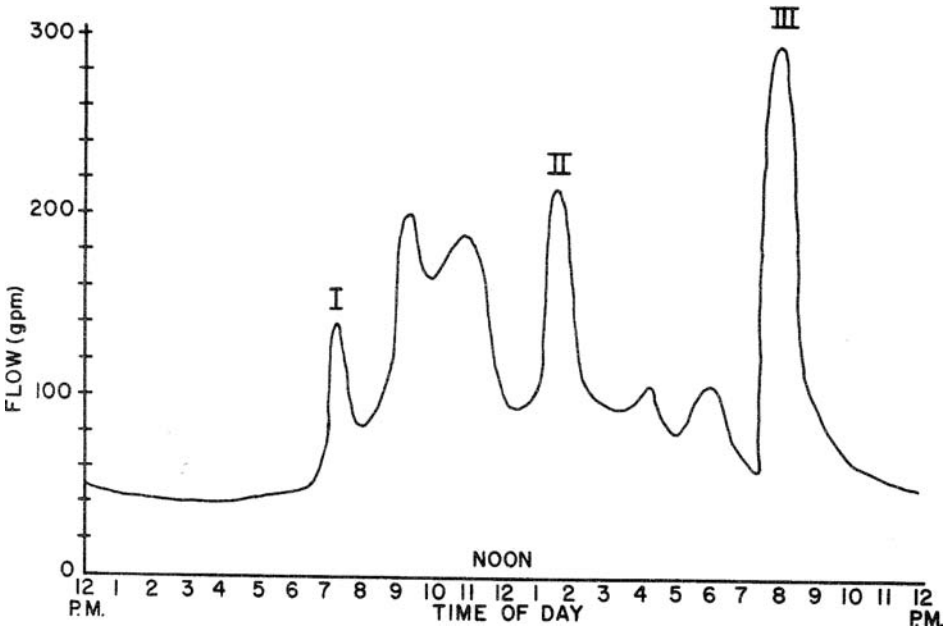

Fig. 7.17. A typical raw sewage flow pattern at Coxsackie sewage treatment plant in November–December 1973.

Table 7.26
Monthly average return sludge characteristics

Month	Year	RSSS (mg/L)	SVI (mL/g)	SS_{SV30} (mg/L)	SRT Based on MLSS (days)	Daily BOD_5 Load on MLVSS (kg/kg)
Aug.	1974	11,400	108	9,300	4.4	0.35
Sept.	1974	11,100	91	11,000	3.7	0.37
Oct.	1974	13,300	93	10,800	5.5	0.26
Nov.	1974	12,000	76	13,200	3.8	0.34
Dec.	1974	13,300	64	15,600	4.8	0.28
Jan.	1975	13,100	64	15,600	6.1	0.26
Feb.	1975	11,300	75	13,300	4.5	0.28
Mar.	1975	12,700	110	9,100	5.2	0.27
Apr.	1975	15,000	73	13,700	6.2	0.23
May	1975	14,200	88	11,400	8.1	0.17
June	1975	14,500	68	14,700	7.0	0.19
July	1975	16,600	59	16,900	9.0	0.17

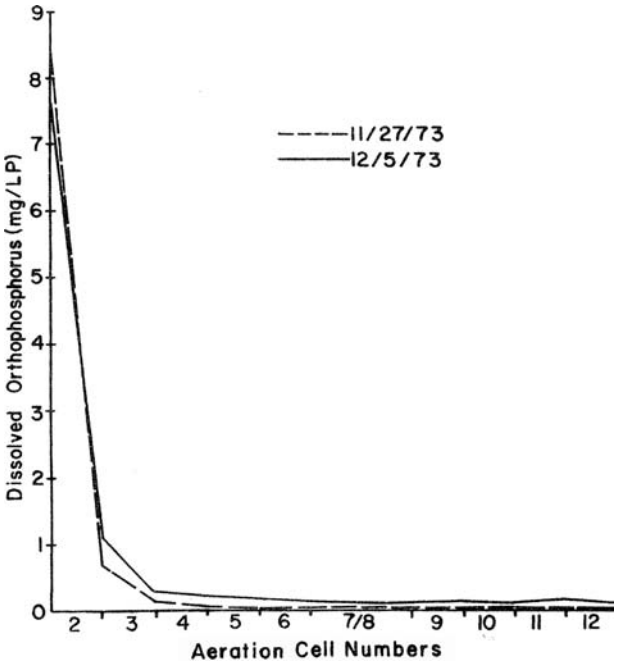

Fig. 7.18. Aeration train dissolved orthophosphorus profiles.

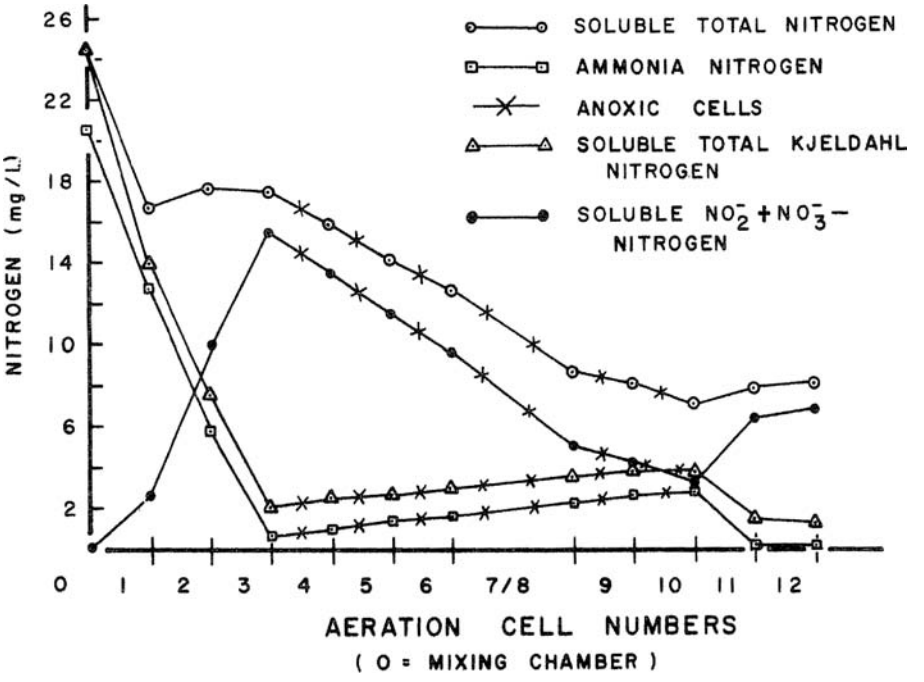

Fig. 7.19. Nitrogen profiles of Coxsackie sewage treatment plant.

8.3.2. Nitrification

Table 7.22 indicates that the amount of ammonia-nitrogen released during ENR was generally much smaller than predicted by stoichiometric theory, which calls for a release of 0.25 mg ammonia nitrogen per mg N gasified, or 25%. To explain the lack of ammonia nitrogen released, one might hypothesize one or more of the following: (a) The adsorbed carbon compounds, or storage carbon not yet assimilated into the biomass, may have been used as the hydrogen donor, which reaction would resemble substrate nitrate respiration and would be accompanied by some sludge production, in turn, resulting in minimal release of ammonia nitrogen. (b) The lack of ammonia nitrogen release may not be real because nitrification may go on simultaneously with denitrification preventing an increase in ammonia-nitrogen concentration, (c) The bacterial biomass may contain less nitrogen than indicated by the formula $C_5H_7O_2N$, which has been stipulated by many researchers (88), and (d) The facultative organisms engaged in nitrate respiration may have used some substrate carbon, perhaps deflocculated matter. The variability of the ammonia nitrogen release observed indicates that more than one of the factors listed above was involved.

8.3.3. Alkalinity Release during Endogenous Nitrate Respiration (ENR)

The theoretical release of alkalinity during ENR is 4.46 mg of alkalinity released per mg nitrate-nitrogen gasified, 3.57 mg of which is attributable to the reduction of nitrate nitrogen and 0.89 mg to the release of ammonia nitrogen by the biomass. The profiles of alkalinity data in Table 7.22 show a lesser release of alkalinity. Three profiles indicate a release near the 3.57 mg level, hinting at the effect of adsorbed carbon respiration; the other two profiles depict a release even below the 3.57 mg level, indicating perhaps nitrification occurring in the anoxic cells.

8.3.4. Alkalinity Destroyed During Nitrification

Table 7.22 indicates that destruction of alkalinity during nitrification was in fair agreement with the stoichiometric prediction of 7.14 mg/mg ammonia-nitrogen oxidized.

8.3.5. Rates of Nitrification

The changing rates of nitrification can best be observed by examining the aerobic portions of the nitrite-nitrate-nitrogen curves. The ammonia-nitrogen curves are not suitable for this purpose because the decrease in ammonia-nitrogen may have been attributable to sorption by the heterotrophic biomass.

The rate of nitrification appears to have been inhibited in the beginning phase of the process. The extent of this inhibition or retardation seems to have depended on the overall nitrification efficiency. At high rates of overall efficiency, the retardation was of short duration, not more than 30 minutes. This retardation seems to be accompanied by intense sorption of ammonia-nitrogen on the part of the heterotrophs. On the other hand, an oxygen deficiency was never noted in any of the aerobic cells. One might therefore theorize that the retardation was attributable to transport difficulties with respect to ammonia nitrogen, the heterotrophic organisms competing with the nitrogen for ammonia-nitrogen and gaining the upper hand

during a short period of the process. Sorption of ammonia-nitrogen was greater than nitrification across the first aerobic cell.

The drastic increase in nitrification rates for the profile shown in Fig. 7.19 reflects the increase of CRT provided for the flow sheet (Fig. 7.16). This increase is in basic agreement with kinetic theory.

The nitrogen profile curves (Fig. 7.19) reflect the mg/L/time rates, not the rates referred to 1 mg/L MLVSS. The nitrogen profile curves are similar to each other. Figure 7.19 only shows a set of nitrogen profiles on June 12, 1975, when the operational conditions were as follows:

1. MLVSS = 4,150 mg/L and MLVSS = 2,600 mg/L
2. Flow-through time = 28 min per cell (except cell 7/8)
3. Mixed liquor temperature = 25 °C
4. Primary settler effluent quality = 36.7 mg/L of TKN; 29.7 mg/L of dissolved TKN; and 26.0 mg/L of ammonia-nitrogen

The nitrification rates appear as the slope of the nitrite-nitrate-nitrogen profiles for the aerobic cells. It is logical to inspect the nitrification rates as they appear in the aeration train profiles for it can be assumed that the active biomass does not change appreciably during one reaction period. Also, when comparing different profiles with each other, the kinetic theory of nitrification in single-sludge activated sludge systems indicates that the mass of nitrogen is independent of MLSS concentration. It is dependent only on the ammonia nitrogen load, environmental conditions, and the growth rate imposed on the MLSS. The average and highest nitrification rates associated with the profiles of Fig. 7.19 are summarized in Table 7.23, on the bases of both mass/volume/time and mass/mass/time.

8.3.6. Rates of Endogenous Nitrate Respiration (Denitrification)

In the aeration train profiles (Fig. 7.19), ENR reaction rates are given in terms of milligram nitrite-nitrate-nitrogen gasified per liter reaction volume versus time and appear as the slope of the nitrite-nitrate-nitrogen profiles for the anoxic cells. In Table 7.23, the rates of denitrification are shown per unit volume of reactor per time and per unit mass VSS per time. The denitrification rates in Table 7.24, obtained by analysis of the mass balance data (32), appear to be higher on the average than the kinetic rates developed in Table 7.23 from the nitrogen profiles of Fig. 7.19. This might be attributable to denitrification occurring in the final settler. A 2 mg/L denitrification effect attributable to the final settler would explain much of the difference encountered.

8.3.7. Phosphorus Removal

Consistently low SS concentrations in the final effluent were established only under the flow sheet of Fig. 7.16 by adding 8 mg/L of Fe^{3+} in the form of $FeCl_3$ to the influent of the primary settler.

Ferric chloride dosage, in terms of mg/L Fe^{3+} and Fe/P molar ratio, is shown in Table 7.20. The Fe/P molar ratio varied between 0.43 and 0.79. Initially ferric chloride was dosed to the effluent of the surge tank. Later the dosing point was relocated to the effluent of the aeration train. This brought about a drastic decrease in $FeCl_3$ requirements and explains the low $FeC1_3$ dosages in later operations. In general, P removal efficiency paralleled SS removal efficiency.

The same unknown wastewater constituent that adversely affected SS removal also affected P removal unfavorably. Efficient P removal by metal addition to the activated sludge process is predicated on the general soundness of the process, as observable in SS removal. When clarification efficiency is impaired because of wastewater characteristics, increased metal dosage generally will not help.

8.4. Solids Management

The relationship between SVI, RSSS, and MLSS concentrations and the effect of BOD_5 load on RSSS concentrations are discussed in this section. The sludge concentration in the space occupied by the settled sludge at the end of a 30-min settled volume test (SS_{SV30}) is related to the SVI as follows:

$$SS_{SV30}=106/(SVI) \tag{14}$$

This equation follows immediately from the definition of the SVI and from geometrical considerations.

SS_{SV30} has been regarded traditionally as approximately equal to the RSSS concentration, provided there is no substantial compaction or dilution of return sludge in the final settler. Such a condition would prevail if the final settler were operated with a rather shallow sludge blanket. SS_{SV30} might be regarded as the maximum safe RSSS concentration that is achievable. In many situations, it will be more practical to operate with a smaller RSSS concentration, i.e., with a higher sludge recycle rate than corresponds to SS_{SV30}. Using Eq. ((14)), Table 7.26 was prepared. The table indicates that SS_{SV30} was approximately equal to RSSS concentration.

Baffle cages were installed in the combined chlorine contact chamber/second phase final settler unit. An average 33% reduction in SS from 5.4 mg/L down to 3.6 mg/L was achieved across the unit.

8.5. Sludge Chlorination Treatment

Data generated during 19 batches of sludge chlorination are summarized in Table 7.27. Additional sludge chlorination treatment results can be found from the literature (32, 83–87). Characteristics of the underflow from the sludge can be found elsewhere (32, 83–87).

A new book edited by Wang, Shammas, and Hung (89) introduces the sludge chlorination process.

The batch size varied somewhat according to the solids concentration of the sludge treated and also according to the sludge storage space available. A typical batch consisted of 34 m^3 (9,000 gal) applied to 111 m^2 (1,200 ft^2) of slow sand bed, resulting in an average sludge slurry dose of 0.31 m (1 ft). However, such a depth was never attained because drainage occurred immediately upon application. At most, approximately 15 cm (6 in.) of chlorinated sludge slurry were seen standing on the bed. Usually, all visible liquid had drained away by the morning following application.

The slurry was treated at a rate of approximately 2.5 L/s (40 gpm). The weighted average of the chlorine dosage was 830 mg/L or 10% of the dry weight of the sludge chlorinated. The chlorine dosage was adjusted to produce a pH of 2.3 to 2.8 in the chlorinated slurry. The pH of the unchlorinated slurry indicates that considerable nitrification had occurred on some batches

Table 7.27
Effect of sludge chlorination on solids concentration, filterability, and pH

		SS			VSS			Filterability			pH	
Date 1975	Batch No.	Before (mg/L)	After (mg/L)	Change (%)	Before (mg/L)	After (mg/L)	Change (%)	Before (mL/30 s)	After (mL/30 s)	Change (%)	Before	After
7/8	1	9190	8680	−6	4950	5020	+1	122	182	+49	6.45	2.6
7/9	2	7400	7240	−2	3780	3680	−3	124	172	+40	6.3	2.65
7/10	3	10,100	9540	−6	5680	5580	−2	20	97	+385	7.0	2.5
7/18	4	8730	8070	−8	5100	5030	−1	158	230	+46	6.7	2.5
7/22	5	10,690	10,420	−3	6190	6470	+5	16	69	+331	6.75	2.7
7/28	6	12,705	11,765	−7	7480	7390	−1	12	67	+458	7.1	2.4
7/29	7	8680	8020	−8	4100	4440	+8	120	162	+35	6.0	2.5
7/31	8	13,440	12,620	−6	7040	7420	+5	14	63	+350	6.4	3.0
8/6	9	9590	9220	−4	5910	6010	+2	10	64	+540	6.6	3.1
8/12	10	9680	9720	–	5640	6080	+8	34	76	+124	6.7	2.75
8/13	11	9780	9460	−3	5550	5600	+9	68	99	+46	6.0	3.0
8/26	12	7300	6920	−5	4860	4820	−1	16	67	+319	6.5	2.6
8/29	13	7360	5920	−20	4720	4130	−13	88	144	+64	5.85	2.4
9/3	14	6260	6710	−7	4100	4890	+19	84	152	+80	6.9	3.2
9/8	15	7050	6330	−10	4220	4030	−5	50	84	+68	6.8	2.4
9/9	16	5440	4950	−9	3320	3340	–	51	82	+61	6.9	2.3
9/11	17	5600	5430	−3	3460	3560	+3	30	77	+157	6.9	2.5
10/6	18	7220	6940	−4	4580	4820	+5	10	83	+730	7.2	2.25
10/7	19	4220	4020	−5	2780	2920	−5	22	24	+10	7.4	2.7
Weighted mean		8180	7760	−5	4790	4900	+2	52	101	+49		

before treatment. During chlorination, samples of unchlorinated and chlorinated sludge slurry were withdrawn and subjected to the tests reflected in Table 7.27. Glass fiber filters were used to determine some of the parameters in the table. There was only a slight reduction in SS attributable to chlorination, approximately 5%. There was an approximate 2% increase in the VSS concentration, probably attributable to experimental error.

The increase in filterability, 49% on the average, was of course the most important change in sludge characteristics brought about by sludge chlorination. The increase of filterability was greater for sludges of a low initial filterability. The weighted average filterability was increased from 52 to 101 mL per 30 seconds. The changes in nutrient concentration were on the average negligible, i.e., under 3 mg/L.

The chlorine residual in the chlorinated sludge was approximately 150 mg/L. A significant increase in the TOC of the sludge supernatant was observed from a range of 11 to 150 mg/L to a range of 46 to 188 mg/L.

The sludge dewatering beds underflow did not impose any significant load on the activated sludge system. The following constituent ranges were determined:

SS = 9 to 76 mg/L
VSS = 2 to 43 mg/L
pH = 5.2 to 6.9 unit
Chlorine residual = 0.1 to 5.3 mg/L
TOC = 39 to 60 mg/L
COD = 142 to 241 mg/L
Alkalinity = 42 to 130 mg/L as $CaCO_3$
NH_4^+-N = 6 to 36 mg/L
NO_2-N + NO_3-N = 1 to 17 mg/L
Dissolved orthophosphorus = 0.4 to 1.8 mg/L

ACKNOWLEDGMENT

Coxsackie Sewage Treatment Plant in New York, USA, is a model single sludge biological treatment plant which was built and operated under the cosponsorship of the New York State Department of Health (NYSDOH), the New York State Department of Environmental Conservation (NYSDEC), and the New York State Department of Correction (NYSDOC). The single sludge plant has been performing successfully to be a role model to the New York municipalities since 1974. Leo J. Hetling (Ph.D., PE), Carl Beer (PE), and Lawrence K. Wang (Ph.D., PE) were Research Director, Project Manager, and Chief Operator (NYSDEC Senior Sanitary Engineer), respectively, in 1973 to 1977. This chapter was written in memory of late Carl Beer, who was a pioneer in the field of single sludge biological systems. Both late Carl Beer and Lawrence K. Wang (senior author of this chapter) received a Kenneth Research Award from the New York Water Environment Association for this environmental engineering contribution.

NOMENCLATURE

b = fractional rate of biomass destruction, decimal fraction or %
C_e = desired effluent nitrate concentration, mg/L
$f_e : f_s$ = ratio of substrate utilized for energy to substrate utilized for synthesis
F/M = food to micro-organisms ratio, g BOD/g MLVSS/day (lb BOD/lb MLVSS/day)
HRT = hydraulic retention, time, hour
I = nitrified internal recycle ratio of recycle rate to plant influent flow rate, decimal fraction or %
MLSS = mixed liquor suspended solids, mg/L
MLVSS = volatile mixed liquor suspended solids, mg/L
Q = plant influent flow rate, m^3/day
Q_1 = anoxic internal recycle rate, m^3/day (MGD or gpm)
RAS = return activated sludge recycle ratio of return sludge rate to plant influent flow rate, decimal fraction or %
SRT = solids or biomass (cells) retention time, day
TKN_{ox} = total mass of oxidizable TKN (nitrates) produced in the aerobic reactor, g/day
v = final fractional biosolids volatility, decimal fraction or %
v_0 = initial fractional biosolids volatility, decimal fraction, or %
WAS = wasted activated sludge, decimal fraction or %
Y = yield factor, mg MLVSS produced/mg BOD_5 removed
θ_c = solids retention time, day
$\theta_c{}^d$ = design solids retention time, day

REFERENCES

1. US EPA, *Nitrogen Control Manual*, U. S. Environmental Protection Agency, Tech. Report # EPA/625/R-93/010, Washington, DC (1993).
2. L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), *Biological Treatment Processes*, Humana Press, Totowa, NJ (2009).
3. N. K. Shammas, *Optimization of Biological Nitrification*, Ph.D. dissertation, Microfilm Publication, University of Michigan, Ann Arbor, MI (1971).
4. L. K. Wang and M. H. S. Wang, Control tests and kinetics of activated sludge process, *Water, Air and Soil Pollution* (Holland), 8, 315–351 (1977).
5. L. K. Wang and D. B. Aulenbach, *BOD and Nutrient Removal by Biological A/O Process Systems*, US Department of Commerce, National Technical Information Service Technical Report #PB88–168430/AS, pp. 12, December (1986).
6. US EPA, *BNR Plant Assessment Interim Data Summary Report*, Office of Wastewater Enforcement and Compliance, Prepared by HydroQual, New Jersey, October (1991).
7. WRC, Theory, Design, and Operation of Nutrient Removal Activated Sludge Processes, Water Research Commission, South Africa (1984).
8. Austgen Biojet Waste Water Systems, Inc., *Design Manual*, Version 1.1, No. 44, May (1991).

9. A. Brenner and Y. Argaman, Effects of feed composition, aerobic volume fraction and recycle rate on nitrogen removal in the single-sludge system, *Water Science And Technology*, (G.B.), 24, 1041 (1990).
10. E. Sotirakou, G. Kladitis, N. Diamantis, and H. P. Grigoropoulou, Ammonia and phosphorus removal in municipal wastewater treatment plant with extended aeration, *The International Journal*, 1, 1, 47–53, March 1999, Source Link http://www.gnest.org/journal/Vol1_No1/sotirakou.pdf, January (2005).
11. A. I. Stamon, Modeling of oxidation ditches using an open channel flow 1-D advection-dispersion equation and ASM1 process description, *Proceedings of the 4th International Symposium on System Analysis and Computing in Water Quality Management*, Quebec, Canada, pp. 313–320, June (1997).
12. B. E. Rittman, and W. E. Langeland, Simultaneous denitrification with nitrification in single-channel oxidation ditches, *Journal of Water Pollution Control Federation*, 57, 4, 300–308 (1985).
13. US EPA, A comparison of oxidation ditch plant to competing processes for secondary and advanced treatment of municipal wastes, EPA/600/2–78/051, Cincinnati, OH (1978).
14. K. Inomae, H. Araki,K. Koya,Y. Awoya, T. Kusuda, and I. Matsuo, Nitrogen removal in an oxidation ditch with intermittent aeration, *Water Science and Technology*, 19 (1987).
15. D. Sen, et al., Process design and operational modifications of oxidation ditches for biological nutrient removal. *Proceedings of 6th IAWPRC Conference*, Prague, Czechoslovakia (1991).
16. L. K. Wang, P. Wang. and N. Clesceri, Groundwater decontamination using sequencing batch process, *Water Treatment*, 10, 2, 121–134 (1995).
17. L. K. Wang, L. Kurylko, and M. H. S. Wang, *Sequencing Batch Liquid Treatment*, US Patent No. 5354458. US Patent & Trademark Office, Washington, DC (1996).
18. D. S. Lee, C. O. Jeon, and J. M. Park, Biological nitrogen removal with enhanced phosphate uptake in a sequencing batch reactor using single sludge system, *Water Research*, 35, 16, 3968–3976 November (2001).
19. G. Bortone, et al. Nitrification, denitrification and biological phosphate removal in sequencing batch reactors treating piggery wastewater, *Water Science and Technology*, (G.B.), 26, 997 (1992).
20. K. L. Norcross, Sequencing batch reactors: an overview, *Proceedings of the 6th Annual Kansas Engineering Conference*, Lawrence, KS (1986).
21. W. M. Shubert, *SCR-Sequencing Batch Reactor*, Aqua-Aerobic Systems, Inc. (1986).
22. K. J. Deeny, J. A. Heidman, W. W. Schuk, K. S. Young, and A. J. Condren, Implementation of sequencing batch reactor technologies in the United States, *The 64th Annual WPCF Conference*, Toronto, Ontario, October (1991).
23. US EPA, *Sequencing Batch Reactors*, Summary report for the Center for Environmental Research Information, EPN625/8–86/011, Cincinnati, OH (1986).
24. R. L. Irvine, D. V. S. Murphy, M. L. Arora, J. L. Copeman, and J. A. Heidman, Analysis of full-scale SBR operation at Grundy Center, Iowa, *Journal of Water Pollution Control Federation*, 59, 3, 132–138 (1987).
25. M. C. Goronszy, CASS, *Cyclic Activated Sludge System: Superior Batch Reactor Technology*, Transenviro, Inc. Aliso Viejo, CA (1990).
26. M. C. Goronszy, Nitrogen removal and sludge bulking control in cyclically operated activated sludge systems, *Proceedings of the 61st Meeting of Ohio Water Pollution Control Association*, Akron, OH (1987).

27. A. C. J. Koot, and J. Zeper, Carrousel, a new type of aeration-system with low organic load, *Water Research*, 6, 401–406 (1972).
28. A. T. VanderGeest, and W. C. Witvoet, Nitrification and denitrification in Carrousel systems, *Progress in Water Technology*, 8, 653 (1977).
29. H. Emori, H. Nakamura, T. Sumino, T. Takeshima, K. Motegi, and K. Tanaka, High rate and compact single sludge pre-denitrification process for retrofit, *Water Science and Technology*, 30, 6, 31–40, (1994). Retrofitting.
30. C. W. Randall, et al., Retrofitting activated sludge plants to improve the removal of nitrogen and phosphorus by biological process, *Water Science and Technology* (G.B.), 22, 275 (1990).
31. B. Andersson and B. Rosen, Upgrading for biological nitrogen removal-some full-scale experience from Sweden, *Water Science and Technology*, (G.B.), 22, 93 (1990).
32. US EPA, *A Study of Nitrate Respiration in the Activated Sludge Process*, US Environmental Protection Agency, EPA-600/2-80-154, Office of Research and Development, Cincinnati, OH, August (1980).
33. US EPA, *Process Design Manual for Nitrogen Control*, US Environmental Protection Agency, Office of Technology Transfer, Washington, DC (1975).
34. G. T. Daigger, *Biological Nutrient Removal*, CH2M-Hill, Denver, Colorado, Paper located on Web Site address: http://bridge.ecn.purdue.edu/∼alleman/w3-class/456/article/daigger-nutrients/daigger.html January (2005).
35. Y. D. Lee, E. B. Shin, Y. S. Choi, H. S. Yoon, H. S. Lee, I. J. Chung and J. S. Na Biological removal of nitrogen and phosphorus from wastewater by a single-sludge reactor, *Environmental Technology*, 18, 10, 975–986, October (1997).
36. R. G. Kavanaugh, and C. W. Randall, Bacterial populations in a biological nutrient removal plant, *Water Science and Technology*, (G.B.), 29, 25 (1994).
37. N. Porges, L. Jasewiez, and S. H. Hoover, Principles of biological oxidation. In: *Biological Treatment of Sewage and Industrial Wastes*, Vol. 1, Mccabe and V. V. Eckenfelder, Jr. (eds.), Reinhold, New York, p. 35 (1956).
38. P. L. McCarty, Energetics and Bacterial Growth. In: *Organic Compounds in Aquatic Environments*, S. U. Faust. and J. V. Hunter (eds.), Marcel Dekker, New York, p. 495 (1971).
39. D. B. Christensen and P. L. McCarty, Multi-process biological treatment model, *Journal of Water Pollution Control Federation*, 67, 2652, (1975).
40. J. L. Barnard, Biological nutrient removal without the addition of chemicals, *Water Research*, 9, 483, May/June (1975).
41. C. Beer, *Liquid Waste Treatment Process*. US Patent No. 3517810. US Patent and Trade Marks Office, Washington, DC, June 30, 1970.
42. K. Wuhrmann, Stickstoff und phosphorelimination, ergebnisse von versuchen im technischen masstab, *Schweizerische Zeitschrift Fur Hydrologie-Swiss Journal Of Hydrology*, 26, 520 (1964).
43. M. H. Christensen, M. H. Christensen, and P. Harremoes, *Biological Denitrification in Water Treatment*, Report No. 72-2, Department of Sanitary Engineering, Technical University of Denmark, Lyngby (1972).
44. R. C. Brenner, Federal government activities in oxygen activated sludge process development. In: *Applications of Commercial Oxygen to Water and Wastewater Systems*, R. E. Speece, and J. F. Malina, Jr. (eds.), Proceedings, Water Resources Symposium No. 6, Center for Research in Water Resources, Austin, TX, p. 135, (1973).
45. Beloit—Passavant Corporation, *Magna Rotor*, Bulletin No. 5300, Birmingham, Alabama.

46. F. J. Ludzak, and M. B. Ettinger, Controlling operation to minimize activated sludge effluent nitrogen, *Journal of Water Pollution Control Federation*, 34, p. 920, September (1962).
47. United Statue Patent No. 3,964,998 issued to Barnard, J. I., June 22 (1976).
48. J. C. Wang, D. B. Aulenbach, and L. K. Wang, Energy models and cost models for water pollution controls – Chapter 29. In: *Clean Production*, K. B. Misra, (ed.), Springer-Verlag, Berlin, Germany, pp. 685–720 (1996).
49. L. K. Wang, J. V. Krougzek, and U. Kounitson, *Case Studies of Cleaner Production and Site Rededication*, United Nations Industrial Development Organization (UNIDO) Vienna, Austria, UNIDO-Registry No. DTT-5-4-95, 136 pp., April (1995).
50. L. K. Wang, and N. C. Pereira, *Handbook of Environmental Engineering, Vol. 3, Biological Treatment Processes*, Humana Press, New Jersey, pp. 520 (1986).
51. C. Beer, L. K. Wang, and L. J. Hetling, Full scale operations of plug flow activated sludge systems, *Journal of New England Water Pollution Control Association*, 9(2) 145–173 (1975).
52. C. Beer and L. K. Wang, Activated sludge systems using nitrate respiration- design considerations, *Journal of Water Pollution Control Federation*, 50, 9, 2120–2131, September (1978).
53. L. K.Wang, Chemistry of nitrification-denitrification process, *Journal of Environmental Science*, 21, 23–28, December (1978).
54. L. K. Wang, Wastewater treatment by biological physicochemical two-stage process system, *Proceedings of the 41st Annual Purdue Industrial Waste Conference*, p. 67 (1987).
55. C. S. Applegate, B. Wilder, and J. R. DeShaw, Total nitrogen removal in a multichemical oxidation system. *Journal of Water Pollution Control Federation*, 52, 3, 568–577 (1980).
56. M. H. Christensen, Denitrification of sewage by alternating process operation, *Progress in Water Technology*, 7, 2, 339–347 (1975).
57. J. L. Barnard, Biological denitrification, *Water Pollution Control* (G.B.) 72, 6, 705–720 (1973).
58. K. Krauth, Practical experiences with nitrification and denitrification processes, *Proceedings of the Japanese-German Workshop on Wastewater and Sludge Treatment*, Tsukuba, October (1982).
59. M. Kuribayashi, Study of nitrified liquor recycled biological nitrification-denitrification processes, *Proceedings of the Japanese-German Workshop on Wastewater and Sludge Treatment*, Tsukuba, October (1982).
60. V. Miyaji, et al., Biological nitrogen removal by step feed process, *Progress in Water Technology*, 12, 6, 193–202 (1980).
61. G. T. Daigger and G. A. Nicholson, Performance of four full-scale nitrifying wastewater treatment plants incorporating selectors, *Journal of Water Pollution Control Federation*, 62, 5, 676–683 (1990).
62. N. K. Shammas, Interactions of temperature, pH and biomass on the nitrification process, *Journal Water Pollution Control Federation*, 58, 1, 52–59 (1986).
63. R. O. Mines Jr., *Designing Single-Sludge Bionutrient Removal Systems*, 2001 World Water and Environmental Resources Conference, Orlando, FL (2001).
64. R. O. Mines, Jr., Design modeling of post-denitrification single-sludge activated sludge process, *Water, Air, and Soil Polluion*, 100, 1–2, 79–88 (1997).
65. P. M. Sutton, K. L. Murphy, and B. E. Jank, Nitrogen control: a basis for design with activated sludge systems, *Progress in Water Technology*, 8, 4/5 (1977).
66. D. Sen, C. W. Randall, and T. J. Grizzard, Critical design and process control features to optimize biological nutrient removal in temperature climates, *Proceedings of 7th IAWPRC Conference*, Washington, DC (1992).

67. R. P. G. Bowker, J. M. Smith, and H. Shah, Assessment of Cost and Effectiveness of Biological Dual Nutrient Removal Technologies in the Chesapeake Bay Drainage Basin, V 1, EPA CBP/TRS 17, Annapolis, MD (1988).
68. P. M. Sutton, and T. R. Bridle, Biological nitrogen control of industrial wastewater, *Water-1980 AICHE Symposium Series*, 177 (1980).
69. J. A. Heidman, I. J. Kugelman, and E. F. Barth, Plug-flow single-stage nitrification-denitrification activated sludge, *The 49th Annual WPCF Conference*, Minneapolis, MN (1976).
70. C. W. Randall, J. L. Barnars, and H. D. Stensel, Design and retrofit of wastewater treatment plants for biological nutrient removal, V 5, *Water Quality Management Library*, Technomic Publishing, Lancaster, PA (1992).
71. P. Llies and D. S. Mavinic, Biological nitrification and denitrification of a simulated high ammonia landfill leachate using 4-stage Bardenpho systems: system startup and acclimation, *Canadian Journal of Civil Engineering*, 28, 1, 85–97 (2001).
72. L. Szpyrkowicz, and S. N. Kaul, Biochemical removal of nitrogen from tannery wastewater: performance and stability of a full-scale plant, *Journal of Chemical Technology and Biotechnology*, 79, 8, 879–888 (2004).
73. D. E. Schwinn, D. F. Storrier, and D. G. Thome, *Full-scale Operation of a Single-Stage Nitrification-Denitrification Plant*, EPA/600/1 2–77/088, Cincinnati, OH (1977).
74. D. E. Schwinn, Final Report on Nitrification-Denitrification Research at the Water pollution Control plant, Town of Barnstable, MA, Stearns & Wheler Engineers and Scientists (1988).
75. N. K. Shammas, An allosteric kinetic model for the nitrification process, *Proceedings of the Tenth Annual Conference of Water Supply Improvement Association*, Honolulu, Hawaii, pp. 1–30, July (1982).
76. IAWPRC, *Simulation of Single-Sludge Processes for Carbon oxidation, Nitrification and Denitrification*, Model developed by the IAWPRC Task Group on Mathematical Modeling for Design and Operation of Biological Wastewater Treatment, Version 1, 1997, Clemson University. Source Link: http://www.ces.clemson.edu/ees/sssp January (2004).
77. G. Esposito, M. Fabbricino, and F. Pirozzi, Four-substrate design model for single-sludge predenitrification system, *Journal of Environmental Engineering*, 129, 5, 394–401, (May 2003).
78. Y. Argaman, G. Papkov, A. Ostfeld, and D. Rubin, Single-sludge nitrogen removal model: Calibration and Verification, *Journal of Environmental Engineering, ASCE*, 125, 7, 608–617, July (1999).
79. G. Boeije, D. Schowanek, and P. Vanrolleghem, Adaptation of the simple treat chemical fate model to single-sludge biological nutrient removal wastewater treatment plants, *Water Science and Technology*, 211–218 (1998).
80. A. C. Di Pinto, Modeling of two stage single sludge system for nitrogen removal, *Environmental Technology*, 11, 509 (1990).
81. C. G. Jih, J. S. Huang, and K. C. Hsieh, Performance evaluation of single-sludge reactor system treating high-strength nitrogen wastewater, *Journal of Hazardous Materials*, 85, 3, 213–27 (2001).
82. BBS, *Single-Sludge Process for Carbon Oxidation, Nitrification and Denitrification*, BBS Engineering. Source Link: http://www.bbsengineering.com/portfolio/alternative.php3, January (2005).
83. C. Beer, L. K. Wang, and L. J. Hetling, Full-scale operation of plug flow activated sludge systems, *Journal of the New England Water Pollution Control Association*, 9(2), September 1975. New York State Department of Environmental Conservation, Albany, New York, Technical Paper No. 42, 1975.

84. C. Beer and L. J. Hetling, *Nitrogen Removal and Phosphorus Precipitation in a Compartmentalized Aeration Tank*, New York State Department of Environmental Conservation, Albany, NY. Technical Paper No. 32 (1974).
85. C. Beer, and L. K. Wang, *Process Design of Single Sludge Activated Sludge Systems Using Nitrate Respiration*. New York State Department of Environmental Conservation, Albany, NY. Technical Paper No. 50 (1977). Presented at the 49th Annual Meeting of the New York Water Pollution Control Association, New York City, January 17–19 (1977).
86. C. Beer, Tests for nitrifying and denitrifying ability of activated sludge. *Bulletin of Environmental Contamination and Technology*, 18, 5, 558 (1977).
87. C. Beer, J. F. Bergenthal, and L. K. Wang, A study of endogenous nitrate respiration of activated sludge. *Proceedings of the 9th Mid-Atlantic Industrial Waste Conference*, Bucknell University, Lewisburg, PA (1977).
88. L. K. Wang, J. H. Tay, V. Ivanov, and Y. T. Hung, (eds.) *Environmental Biotechnology*. Humana Press, Totowa, NJ (2009).
89. L. K. Wang, N. K. Shammas, and Y. T. Hung, (eds.) *Biosolids Treatment Processes*. The Humana Press, Totowa, NJ, 820 p. (2007).
90. L. K.Wang, Y. T.Hung, H. H. Lo, and C. Yapijakis,(eds.), *Handbook of Industrial and Hazardous Wastes Treatment*, CRC Press/Marcel Dekker, New York, pp. 415–467 (2004).
91. P. Nawghare, N. N. Rao, and R. Bejankiwar, Szyprkowicz. Treatment of phosphoric acid plant wastewater using Fenton's reagent and coagulants, *Journal of Environmental Science and Health*, A 36, 10, 2011 (2001).
92. Wang, L. K. *Poly Iron Chloride and Poly Aluminum Chloride*, Technical Report LIWT/l-2002/252; Lenox Institute of Water Technology, Lenox, MA, 26 p. (2002).
93. Y. T. Hung, S. Gubba, H. H. Lo, L. K. Wang, C. Yapijakis, and N. K. Shammas, Application of wetland for wastewater treatment. *OCEESA Journal*, 20(1), 41–46 (2003).
94. L. Johansson, Industrial by-products and natural substrata as phosphorus sorbents, *Environmental Technology*, 20, 3, 309 (1999).
95. J. W. McGrath, and J. P. Quinn, *Phosphate Removal: A Novel Approach*, School of Biology and Biochemistry, Queens' University Belfast, Ireland (2002); http://www.qub.ac/uk/envres/EarthAirwater/phosphate-removal.htm
96. B.S. Akin and A. Ugurlu, Enhanced phosphorus removal by glucose fed sequencing batch reactor. *Journal of Environmental Science and Health*, A 36, 9, 1757 (2001).
97. CEEP Phosphates, a sustainable future in recycling. *Centre Europeen d' Polyphosphates*, (1999).
98. A. Giesen, Eliminate sludge. *Industrial Wastewater*, 6 (1998).
99. D. F. Martin, P. M. Dooris, and D. Sumpter, Environmental impacts of phosphogypsum vs. borrow pits in roadfill construction. *Journal of Environmental Science and Health*, A 36, 10, 1975 (2001).
100. N. Priyantha and S. Pereira, Removal of phosphate, sulfate, and colored substances in wastewater effluents using *Feldspar. Water Research and Management*, 14, 6, 417 (2000).
101. M. Krofta and L. K. Wang, Improved biological treatment with a secondary flotation clarifier. *Civil Engng for Practicing and Design Engineers*, 2, 307–324 (1983).
102. L. K. Wang, *An Emerging Technology for Phosphorus Removal from Wastewaters*, Technical Report LIWT/l-2002/253; Lenox Institute of Water Technology: Lenox, MA, 18 p. (2002).
103. L. K. Wang, Laboratory simulation of water and wastewater treatment processes. *Water Treatment*, Vol. 10, 261–282 (1995).
104. Anonymous. Flotation equipment. *Environmental Protection*, 14, 2, 137–138 (2003).

105. L. K. Wang, M. H. S. Wang, and J. Wang, *Preliminary Design Report of a 10-MGD Deep Shaft-Flotation Plant for the City of Bangor*, PB88–200605/AS; US Department of Commerce, National Technical Information Service: Springfield, VA, 171 p. (1987).
106. State of Florida. I*ndustrial Wastewater Program—Phosphate Industry*. State of Florida, Dept. of Environmental Protection, Miami, FL, (February 2004). www.dep.state.fl.us.
107. L. K. Wang, Y. T. Hung, and N. K. Shammas, (eds.) *Physicochemical Treatment Processes*, Humana Press, Totowa, NJ, 723 p. (2005).

8
Selection and Design of Nitrogen Removal Processes

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CONTENTS

Abstract The characteristics of the wastewater to be treated and site constraints will affect treatment performance and thus the selection of an effective process. The nature of the existing facilities will have an effect on the process selection when upgrading for nitrogen removal, especially when attempting to make maximum use of the existing facilities to reduce costs. Usually, a single-sludge system can be more easily retrofitted into an existing activated sludge plant than can a separate-stage system. In addition to the discussion of the factors that affect process selection, the chapter covers costs, design considerations, process design, and design examples.

Key Words Design and costs • nitrogen removal • performance • process selection • single-sludge system.

1. FACTORS THAT AFFECT PROCESS SELECTION

1.1. Wastewater Characteristics

The characteristics of the wastewater to be treated will affect treatment performance and thus the selection of an effective process. Of primary concern to single-sludge nitrification-denitrification systems is the ratio of BOD_5 to TKN. Organic carbon is required by the denitrifying organisms. The BOD_5:TKN is an indication of the supply of necessary carbon,

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_8 © Humana Press, New York, NY 2009

with a high ratio favoring denitrification. Also significant is the presence of readily available organic carbon in terms of soluble BOD_5($SBOD_5$), which is indicated by $SBOD_5$:BOD_5. A high proportion of soluble and readily degradable BOD_5 would favor denitrification and improve reaction rates (1).

Wastewater temperature is a critical parameter because it affects the growth rate of nitrifiers and thus the design θ_c and also the rate of denitrification. The availability of adequate capacity in an existing wastewater treatment plant is therefore significantly affected by the design temperature (2–5).

The pH of wastewater also affects nitrification and denitrification rates. The optimum pH range for nitrification is generally accepted to be 6.5 to 8.5. For denitrification the optimum pH is 7.0 to 8.0 (1–6). Because nitrification consumes alkalinity, the natural bicarbonate alkalinity in a wastewater is of concern. If the alkalinity remaining after ammonia oxidation is $<50\,mg/L$ (as $CaCO_3$) then provision must be made to supplement the alkalinity. Single-sludge nitrification-denitrification systems will have alkalinity returned to the process as a result of the denitrification reaction. Approximately 50% of the alkalinity lost through nitrification can be regained during denitrification in a single-sludge system, if all nitrates are denitrified.

Variability in flows and loads will negatively affect process performance. Wastewater treatment facilities with highly variable flows should consider the addition of flow equalization. An equalization basin can also dampen peak loads caused by internal recycle flows such as digester supernatant returns and dewatering operations. Because of the concern for the BOD_5:TKN as described previously, the possibility of BOD_5 and TKN peaking at different times should also be investigated.

Wastewaters that include a significant contribution from industry should be investigated for substances that may be inhibiting to the nitrification process. A separate-stage nitrification system affords some protection to the nitrifiers by providing buffering capacity in the first-stage carbonaceous BOD removal process (4). Inhibitory compounds can significantly reduce the rate of nitrification. The addition of powdered activated carbon may enhance nitrification rates in these cases.

Very high ammonia concentration (in the NH_3 form) can be toxic to nitrifiers. The amount of ammonia present as NH_3 is dependent on both pH and wastewater temperature, with its relative concentration increasing as the pH and wastewater temperature increases (1).

Collection systems that suffer from a high degree of infiltration/inflow or that contain combined sewers will produce a dilute wastewater. Such wastewater will exhibit lower rates of denitrification because of the lower concentration of organic carbon.

Wastewaters that include septage may contain a relatively higher fraction of refractory TKN. This form of nitrogen is resistant to treatment and may require long θ_c's to achieve even partial oxidation.

1.2. Site Constraints

If the space available for plant upgrade or expansion is limited, then single-sludge systems, which do not require intermediate clarifiers, should be considered (1, 7–9). Although the reactor volume required for single-sludge nitrification-denitrification will be greater than

that required for a separate-sludge system, the total combined volume including reactor and settling capacity may be less for a single-sludge system.

In situations where sufficient space is not available to expand immediately adjacent to the existing activated sludge tankage, it may be necessary to implement a separate-stage system and use the existing tankage for carbonaceous BOD removal only. The second stage can be built elsewhere, though pumping may be required. Alternatively, it may be possible to divert some of the plant influent to new tankage, thus allowing nitrification-denitrification to be incorporated into existing tankage. Plans for upgrade and expansion should also consider future requirements so that process selection and site planning do not preclude treatment alternatives in the future.

Where space is a concern, consideration should be given to higher rate processes such as the VIP process (Virginia Initiative Plant) (10–12). This process, which was developed to optimize biological phosphorus removal, operates at a total θ_c of 5 to 10 days under warm weather conditions compared to the modified Bardenpho or UCT (University of Capetown) (10–12) processes, which typically operate at θ_c's of 10 to 25 days. Also, the VIP process promotes higher rates by creating multiple compartments instead of single complete mix reactors for the anaerobic, anoxic, and aerobic zones. This approximates a plug flow reactor, which increases the substrate concentration in the initial compartments of each zone, thereby increasing the reaction rate. The overall result is that TN removal can be achieved in a smaller volume and in less area. The higher rates associated with separate-stage denitrification as compared to denitrification in a single-sludge system, can also reduce land requirements.

SBR (sequencing batch reactor) (13, 14) systems offer the possibility of a high degree of treatment in a relatively small space. Because settling occurs in the aeration basin, there is no need for a final settling tank. Also, this allows operation at relatively high MLSS concentrations, which can decrease volume requirements.

There are also design considerations that can reduce land requirements. These include the use of common wall construction with rectangular settling tanks, the use of deeper aeration tanks with fine bubble aeration, and the use of methanol to increase denitrification rates.

1.3. Existing Facilities

The nature of the existing facilities will have an effect on the process selection when upgrading for nitrogen removal, especially when attempting to make maximum use of the existing facilities to reduce costs. Usually, a single-sludge system can be more easily retrofitted into an existing activated sludge plant than can a separate-stage system. This is particularly true if there is sufficient excess capacity available to allow the anoxic zone or zones to be incorporated without building additional tankage. A separate-stage system will require intermediate clarifiers and process tanks and may require an intermediate pumping station (1).

Effluent limits will govern process selection, but where a single-sludge, single anoxic zone system is applicable, a high-rate design can be more easily retrofitted where excess capacity is limited. Additional baffling within zones to approximate plugflow kinetics can improve nitrification and denitrification efficiency. The VIP process exploits this type of design.

There may be occasions where the configuration of the existing aeration basin does not allow the basin to be divided into the proper size compartments or the baffles to be located where desired to create separate aerobic/anoxic zones. For example, a mechanically aerated basin will typically be divided into compartments with the aerator at the center, and it may be difficult to install a new baffle at any location other than where the aeration compartments are already divided. One possible solution is to dedicate the first compartment to an anoxic condition but then employ cyclical aeration in the second compartment. The total effective anoxic volume may then be varied by changing the on/off time of the aerator in the second basin. These schemes would require internal recycle of nitrates to the anoxic zone.

Existing facilities may limit the level of MLSS (mixed liquor suspended solids) that can be carried in a system either because of the loadings on the final clarifier or the capacity of the return sludge pumps and piping systems. This could limit the performance of a system that must nitrify year-round, if in winter the system cannot carry enough MLSS to allow nitrification to continue. If the capacity of the clarifiers or the return sludge system cannot be improved, then a possible solution is to install fixed-film media in the aerobic section of the tank. This would increase the nitrifier population and the effective θ_c of the nitrifiers without burdening the clarifiers. This technology has been applied in Japan and Europe, but is relatively new to the United States.

Nitrification and denitrification reaction rates are temperature dependent. If the existing facilities do not provide enough tank volume to nitrify and denitrify year-round in a single-sludge system, then it may still be cost effective to provide a single-sludge system and add a separate-stage denitrification step. During the summer, when reaction rates are greater, the single-sludge system may be adequate to denitrify and thus may provide the cost savings associated with the return of alkalinity and the use of nitrates to oxidize organics. During the winter, when more of the existing tankage is required to nitrify, the separate-stage system could be operated for denitrification (1).

2. COSTS

2.1. Capital Cost

The cost for upgrading existing wastewater treatment facilities or adding new facilities for biological nutrient removal is site specific and varies considerably. Such factors as the actual BOD and TKN loads, the nature of the existing facilities, site conditions and degree of new versus retrofit facilities will have major impacts on the design and cost of the facilities and it is difficult to provide meaningful generalizations relating cost to design flow (1, 15–17).

In many cases, single-sludge systems will have a lower capital cost than separate-sludge systems, primarily because a single-sludge system does not require intermediate clarifiers. Estimates based on studies in the literature indicate that a separate-sludge system can typically cost 15% to 20% more than a single-sludge system. General guidelines for costs of various components of a biological nutrient removal (BNR) system have been published (18). These guidelines provide approximate costs based on a range of assumptions and are not intended to replace a detailed cost estimate for a specific installation.

2.2. Operational Cost

Single-sludge systems offer several potential advantages over separate-stage systems that can reduce their operating costs. Aeration requirements are reduced in a single-sludge system when wastewater is used as the carbon source for denitrification in the anoxic zone. Nitrates replace oxygen as the electron acceptor in oxidizing carbonaceous BOD in the denitrification reaction. The net affect is to reduce the aeration required for BOD removal by as much as 25%. This is partially offset by extra mixing energy required by anoxic reactors and larger aeration tanks.

In addition, the use of wastewater as the carbon source can eliminate the need for methanol addition as in separate-stage denitrification systems. Methanol addition adds a significant operating cost and is a hazardous material to handle. This is potentially or partially offset by the high capital and operation costs of mixed liquor recycle.

The need for supplemental alkalinity is also reduced or eliminated in a single-sludge system. Approximately one-half of the alkalinity lost during the nitrification reaction can be recovered during the denitrification reaction; however, this is not of significance in alkaline or adequately buffered waters.

Single-sludge systems have been shown to produce less sludge than a separate-sludge system. The relatively long θ_c in single-sludge systems results in increased endogenous respiration and thus less excess biomass to be wasted. Also, BOD oxidation by nitrates under anoxic conditions minimizes heterotrophic biomass production. Therefore, when evaluating single-sludge systems for nitrogen removal, the impact on sludge production should be considered.

Treatment systems with permits requiring both phosphorus and nitrogen removal may use biological phosphorus removal rather than just chemical precipitation to reduce operating costs. Chemical precipitation for phosphorus removal may reduce the amount of alkalinity present in the wastewater to the point where supplemental alkalinity is required to avoid pH depression. By incorporating biological phosphorus removal it may be possible to limit the loss of alkalinity to the point where chemical supplementation is not required.

3. DESIGN CONSIDERATIONS

3.1. Primary Settling

The use of primary settling tanks will provide the usual benefits associated with such systems including the reduction of rag accumulations on aeration equipment; the reduction of nonbiological floatables in the aeration tanks and final settling tanks; and process improvements related to the capture of solids from return flows such as digester supernatant and thickener overflow, and from septage discharges (1). The removal of BOD in the primaries will reduce the volume required by the biological reactor for carbonaceous BOD removal and nitrification. However, there are additional factors to consider with a single-sludge system for nitrification and denitrification. Primary settling will also reduce the BOD_5:TKN, which may reduce the rate of denitrification that can be achieved. This may not be a significant problem if a large fraction of the total BOD is soluble, in which case, the removal of particulate BOD

may not adversely affect denitrification. A BOD_5:TKN > 5 favors denitrification. If primary settling is contemplated, a short settling period or the use of fine screens should be considered; alternatively, provision should be made to bypass a portion of the raw wastewater around the primary settling tanks to increase available carbon for denitrification (19, 20).

3.2. Aeration Systems

The aeration system must be sized to handle the increased oxygen demand imposed by nitrification and must be capable of delivering the total amount of oxygen required for complete carbonaceous BOD removal and nitrification under peak loading conditions and changing seasonal conditions. Plug flow designs must consider the greater oxygen demand at the head end of the tank. Additional cost savings can be obtained by installing a DO (dissolved oxygen) monitoring/aeration control system to vary the blower output in response to the oxygen demand.

Seasonal and diurnal variations in total oxygen requirements can cover a large range. Diffused air systems, with the turn-down capability inherent in blower equipment and the ability to taper the aeration capacity, can take advantage of these variations and provide savings in operating costs. Fine pore aeration systems are recommended over coarse bubble because of their increased oxygen transfer efficiency, but with that comes an increased potential for fouling. Where cyclical aeration is used, ceramic fine bubble diffusers should not be employed; flexible membrane-type diffusers have been used with cyclical start/stop operation in small systems and should be considered. For cyclical systems, electrically operated butterfly valves should be provided on air headers to allow cycling of the air supply to various tank compartments.

Mechanical surface aerators with conservative service factors require less maintenance but do not have the same degree of turn-down capability as diffused air systems. Also important in northern climates is the tendency of mechanical aerators to increase heat loss. Mechanical aerators are frequently used for cyclical nitrogen removal systems because they can be easily cycled on and off at set intervals using programmable timers. Aerator cycles may be staggered to avoid high-ampere draws upon aerator startup, or connected to a variable frequency drive. Timers should be adjustable to allow each on- or off-cycle to vary over a range of 30 minutes to a few hours as well as allow various cycle patterns at different times of the day and different days of the week. In existing plants that are being operationally modified for nutrient removal, mechanical aerators can be converted to mixers for use in an anoxic zone. One oxidation ditch technology uses variable-speed rotors combined with weir level control to yield a highly flexible range of aeration and mixing conditions, which provides conditions that transcend some of the above issues (1).

Submerged turbine aerators provide some of the benefits of both the diffused air and mechanical surface aeration system. They do offer some turn-down capability, at least in regard to the air supplied to the diffuser. An anoxic zone can be easily created by shutting the air completely off, in which case the turbine would serve as a mixer.

Dissolved oxygen monitoring should be considered for any system that incorporates aerobic and anoxic zones. DO information is critical to optimizing system performance. This is especially true when operating a plant in the CNR (cyclical nitrogen removal) (21) mode or

when attempting to operationally modify an oxidation ditch for BNR. Automated DO control should be considered in most systems to save energy and to control the process. For cyclical aeration, the DO level during the aerobic phase should be maintained at 1 to 2 mg/L.

3.3. Mixers

Submerged propeller mixers or turbine mixers are typically used to maintain the MLSS in suspension in the anoxic zone. The location of the mixer(s) is critical to proper operation and the manufacturer must be consulted regarding this matter. The objective is to provide mixing energy without turbulence, which would entrain air, and to avoid dead spots, which could become anaerobic. Mixers are desirable during the anoxic phase with cyclical aeration but are not mandatory if the off-cycle is short. Consideration should be given to aerator designs that can provide mixing during the anoxic cycle, such as submerged jets (1).

3.4. Recycle Pumping

The pumping of nitrified mixed liquor from an aeration zone to an anoxic zone to recycle nitrates for denitrification is typically required. This will often require pumping from one end of the aeration tank to the other, over a tank wall or flow channel, or through the aerator basin wall. In these cases, the water level in the aerobic and anoxic zones is approximately the same and the system head will normally be low. However, pumping of large volumes, as much as four times the plant influent flow may be required. Larger pumping volumes are impractical because the marginal increase in nitrogen removal via internal recycle decreases significantly for recycle rates >400% of the influent flow. The rate of nitrate recycle controls the denitrification process in the first anoxic zone and establishes the maximum efficiency achievable assuming the wastewater organic content is sufficient. Multiple smaller pumps should be provided in lieu of a few large pumps to control the recycle rate as changing conditions dictate to optimize the process. DO concentrations in recycle streams should be kept to a minimum (1).

3.5. Reactor Design

Aerobic and anoxic zones should be designed to allow for flexibility in operation to optimize the various processes by the use of channels, piping, gates, and valves such that alternate feed points or tanks and compartments can be used for influent, internal recycles, and return sludge. Control of DO levels, solids inventory, recycle rates, sludge blanket levels, and tankage in service is necessary to optimize virtually all of the processes given the impact of changes in diurnal loadings, seasonal loadings, and temperature changes (22–25).

Submerged baffles are desirable to divide the anoxic zones into compartments operated in series to simulate a plug flow type configuration. Multiple compartments in the nitrification zone may be desirable to avoid short-circuiting of ammonia and to ensure that the internal recycle flow to the anoxic zone has been fully nitrified. For cyclical or multiple anoxic zone nitrogen removal processes, the ability to step-feed influent flow to downstream compartments may be desirable to provide wastewater as a carbon source during denitrification in lieu of an internal recycle.

3.6. Secondary Settling

Biological nutrient removal systems are susceptible to the same operational problems experienced with typical activated sludge systems and may be plagued by some additional problems attributable to the presence of anaerobic and anoxic zones. Bulking sludge may occur with the growth of filamentous organisms. One possible cause of bulking sludge is a condition of low dissolved oxygen. This situation may occur if close control of the aeration system is not maintained during periods when oxygen demand is increasing such as in plants that transition seasonally from carbonaceous BOD removal only to operation with nitrification. The use of chlorine to control the growth of filamentous organisms can be effective. However, this practice may be harmful to the performance of plants that also incorporate biological phosphorus removal, because the chlorine can also oxidize the soluble organic substrates required for efficient biological uptake of phosphorus. Excessive anoxic retention periods may also promote bulking sludge. The total anoxic period should not exceed the time required for denitrification of the nitrate mass returned via the recycles (26).

The nuisance organism *Microthrix parvicella*, which produces scum and is difficult to eliminate with anoxic selectors only, has been reported at biological nutrient removal plants. Design of BNR facilities must assume that foam and scum will occur and provide adequate facilities for the collection and disposal of scum and floating solids from clarifiers.

The addition of an aerobic stabilization zone before the final settling tank has been reported to improve settling performance. Improved performance is likely the result of increasing the DO level in the influent to the final settling tank, thus preventing denitrification. Nitrogen gas, produced by denitrification, attaches to sludge particles causing them to rise. Also, the additional aerobic detention can prevent denitrification by oxidizing remaining wastewater organic matter or any remaining methanol if it is used in a postdenitrification stage. This would eliminate a carbon source for denitrification. However, this approach has a potential negative impact on systems that recycle RAS (return activated sludge) to the anoxic zone. The RAS is more likely to have a level of DO that will tend to decrease the denitrification rate.

Another possible solution for systems that may be plagued by rising sludge is to provide rapid sludge removal equipment, such as vacuum collector final settling tanks.

3.7. Selectors

Several researchers have observed poor sludge settling characteristics in nitrogen removal processes (26–28). Nuisance filamentous organisms *M. parvicella, Sphaerotilus natans, and Nocardia* have been identified in bulking sludge samples; and their presence has been determined to induce bulking conditions. The organism most often identified in bulking sludges is *M. parvicella*, which has been characterized as a low F/M micro-organism. Low F/M organisms exhibit a higher growth rate at low substrate levels (26). Consequently, they will proliferate at low F/M, suppressing the growth of floc forming bacteria. Conversely, at high substrate concentrations, floc-forming zoogleal organisms maintain higher growth rates and are able to out-compete filamentous organisms. Thus, bulking sludges attributable to *M. parvicella* can be suppressed by providing a zone with high substrate loading conditions.

Other causative factors of bulking sludges include anoxic mixing sequences (28), low BOD_5:N and BOD_5:P (29), and low DO levels. As a result of the variety of relationships recorded that were determined to cause bulking, no single process variable has been acknowledged as a process control parameter (30). Mixing return sludge with influent wastewater in one or more in-series contact chambers for a short duration before directing the stream into a complete mix basin has been suggested (31–33); the pre-react chamber described was termed a "selector" because it affects the selection of nonfilamentous organisms. Biomass grown under a substrate gradient loading condition has been observed to control sludge bulking in both aerobic and anaerobic selectors (28). This observation has been confirmed at several full-scale plants (34).

The following have been recommended for effective selector design (30):

1. Selector should be designed with a sharp soluble organic substrate gradient.
2. Substrate leakage from the selector should be minimized. The selector should be designed to remove more than 90% of soluble substrate.
3. Microbial activity (determined from the substrate uptake rate) should be maintained as high as possible.

Although aerobic, anaerobic, and anoxic selectors have all been found to control bulking effectively, the type of aeration was determined to influence selector performance (34). Mechanical aerators were hypothesized to produce a DO gradient in the aeration basin, which, in conjunction with an aerobic selector, resulted in a poorly settling sludge. An anaerobic plug flow selector, however, was effective in controlling sludge bulking when placed ahead of aeration basins with surface mechanical aerators. Whether effective selector performance with mechanical surface aerators was attributable to the plug flow regime or the anaerobic conditions, or both, the flow regime and anaerobic condition in concert, could not be determined. Although both aerobic and anaerobic selectors can control bulking, anaerobic selectors can also provide the benefit of phosphorus removal without requiring additional aeration capacity.

4. PROCESS DESIGN

4.1. Introduction

The following design features are some of the factors to be considered during a facility sizing and design phases:

***Internal Recycle Rate*:** The amount of denitrification in systems with internal recycle is controlled by the rate of recycle to the anoxic zone. There is a practical limit, or point of diminishing returns, even when the influent BOD_5:TKN is adequate (at least 3:1) such that waste-water carbon is the carbon source. Internal recycle pumps should generally be sized to provide an upper limit of three to four times the influent flow rate, except in unusual cases.

***DO Control*:** Automatic DO control for the aerobic zones is desirable to reduce energy consumption and to prevent high DO levels in the internal recycle to the anoxic zone, which could adversely affect the denitrification process. DO levels in the anoxic zone should be

less than or equal to 0.3 mg/L at all times. A tapered aeration system is appropriate for a plug flow configuration.

***RAS and WAS Pumping Rates*:** Variable-speed pumps or flow-control valve arrangements should be provided to control and vary pumping rates to adjust to changes in influent loadings, reactor temperatures, and taking tanks in and out of service.

***Internal Recycle Pumping*:** Pumps should be located at the end of the nitrification zone where they will minimize DO levels in the recycle flow. Recycle flow should be returned to the anoxic zone via piping and should be submerged at the point of discharge. Multiple pumps are desirable to vary the internal recycle rate depending on changing conditions.

***Multiple Basins*:** Multiple basins should be provided to allow taking basins out of service during warm weather or low loading periods; therefore, flexibility in piping, valves, gates, and channels is desirable to operate the system as needed.

4.2. Summary of Design Procedures

The following is an outline of procedures used in designing single-sludge nitrification-denitrification systems (1):

1. Determine influent characteristics, effluent limitations, time basis of limits (e.g., monthly, weekly), peaking factors and design temperature based on weekly or monthly minimum average temperature for the time period that the nitrogen limits are in effect.
2. Prepare mass balances for the entire plant for the annual average, maximum monthly, and maximum weekly or peak day conditions that could affect the design calculations. The mass balances should reflect the impact of all recycle streams and any intermittent discharges, such as septage or landfill leachate.
3. Calculate the level of treatment required for denitrification and TN removal. All systems generally will be designed to achieve complete nitrification. Select type(s) of single-sludge process configurations required to achieve the desired level of treatment to meet the effluent limits with a margin of safety.
4. Calculate the volume and MLSS required for the nitrification zone based on aerobic design θ_c^d and controlling conditions at the final clarifier.
5. Determine the size of the first anoxic zone based on the degree of denitrification required and/or achievable with various internal recycle rates, where applicable. RAS rates should be included with internal recycle rates for single anoxic zone systems with predenitrification. Select the denitrification rate based on the carbon source to be used and adjust for temperature and peaking factors or maximum design loading. Where feasible, denitrification rate studies should be conducted before selecting the denitrification rate used in design. Wastewater typically would serve as the carbon source where the influent (feed to secondary system) BOD_5:TKN is at least 3:1. A trial-and-error solution might be required to size the anoxic zone, because the denitrification rate is dependent on the anoxic F/M ratio (availability of COD as the carbon source) (5, 19, 20, 24, 26, 35, 36).
6. Size the second anoxic zone based on the nitrate loading that was not denitrified in the first anoxic zone, the additional denitrification required, and the selected denitrification rate using an endogenous carbon source. The rate should be adjusted for temperature and maximum loading used for design. Alternatively, methanol can be used in the second anoxic zone (5, 19, 20, 24, 26, 35, 36).

7. Size the postaeration zone to achieve a residual DO level of 1 to 2 mg/L before the secondary clarifier.
8. Determine WAS (waste activated sludge) and RAS pumping requirements to cover the full range of possible conditions (37).
9. Calculate aeration requirements for nitrification and mixing requirements for the anoxic zone. Generally, the aeration system should be sized for nitrification without the oxygen demand savings from denitrification if sized on a maximum monthly basis. Peak day and short-term peak demands should also be considered in determining total aeration capacity. The minimum oxygen demand condition should be determined as well to ensure that the aeration system has adequate turn-down capability to control DO levels as desired.
10. Determine alkalinity requirements to ensure a residual of at least 50 mg/L as $CaCO_3$. Alkalinity produced by denitrification should be included in the calculations.
11. Prepare final mass balance to check sizing of unit processes and redo calculations as necessary.

5. DESIGN EXAMPLES

5.1. Introduction

The following process design examples illustrate sizing calculations for two different plant scenarios with two different sets of effluent limitations (1). Both plants in the scenarios are activated sludge plants and have an average daily design flow of 220 L/s (5 MGD). Plant A (Appendices A and B) does not have either primary settling or separate digestion. Plant B (Appendices A and C) has primary settling tanks and anaerobic digesters followed by mechanical sludge dewatering. For each plant, two sets of effluent limitations are imposed. One set consists of secondary treatment standards with nitrogen removal on a seasonal basis to meet a total nitrogen (TN) limit of 10 mg/L. The other set requires advanced waste treatment for BOD_5 and SS with a TN limit of 5 mg/L on a year-round basis; effluent filtration is provided. For the more stringent limitations, the impact on nitrogen removal of imposing an additional limit of 1.0 mg/L total phosphorus (TP) is also considered for that design example. Table 8.1 summarizes the two sets of effluent limits.

Table 8.1
Effluent limits for treatment plants

		Effluent limits, mg/L	
		30-day	7-day
Effluent 1 (seasonal)	TN	10	15
	NH_4^+-N	2	3
	$CBOD_5$	30	45
	TSS	30	45
Effluent 2 (year round)	TN	5	7.5
	NH_4^+-N	2	3
	$CBOD_5$	10	15
	TSS	10	15

Source: US EPA.

The following examples illustrate process design and sizing for single-sludge nitrification-denitrification systems, which primarily involves sizing of the nitrification and anoxic zone(s) or phases in the aeration tanks to achieve nitrogen removal. The design examples also illustrate other design features, requirements, and/or impacts on support systems. Specifically, the examples identify reactor size, typical reactor configuration, aerator/mixing requirements, waste and return sludge requirements and internal recycle rates. All calculations shown are based on designing for 15°C (59°F) water temperature. The sizing results for designs at 10°C (50°F) and 20°C (68°F) are also summarized for comparison. The plant scenarios illustrated are not intended to suggest optimum approaches but to demonstrate calculations under different conditions.

For the four plant and effluent scenarios selected as design examples, three of the major classifications for nitrogen removal are used to illustrate process design for that type of system. Each type of system used is intended to be generic and does not reflect sizing techniques for any particular system offered by manufacturers either proprietary or nonproprietary. The following systems are used for the design examples (1):

Design Example 1. Single Anoxic Zone—Plant B (complex plant) and less stringent limits.
Design Example 2. Dual Anoxic Zones—Plant B (complex plant) with more stringent limits.
Design Example 3. Multiple Anoxic Phases (cyclical)—Plant A (simple plant) with less stringent limits.
Design Example 4. Dual Anoxic Zones—Plant A (simple plant) with more stringent limits.

5.2. *Design Example 1: Plant B with Less Stringent Limits*

From mass balances (Appendix D), the inlet wastewater characteristics (primary effluent with recycles) and secondary effluent characteristics for the more complex plant without nitrification-denitrification are as shown in Table 8.2.

To meet seasonal limits of 2 mg/L NH_4^+-N and 10 mg/L TN, design for 1.0 mg/L equivalent NH_4^+-N and 8 mg/L equivalent TN at maximum monthly loadings. Secondary effluent Org-N of approximately 2.0 mg/L equivalents represents the nonbiodegradable fraction of soluble TKN and nitrogen associated with effluent VSS.

With an influent raw TKN concentration of 30 mg/L and effluent limits of 2 mg/L NH_4^+-N and 10 mg/L TN, essentially complete nitrification and a minimum of 67% TN removal efficiency are required. A single anoxic zone-type process, such as the MLE (Modified Ludzack-Ettinge; refer to Chapter 7) process, is adequate to meet these limits. The plant configuration is as shown in Appendix C, for the more complex Plant B using a single aerobic zone preceded by an anoxic zone for nitrogen removal with internal recycle, mixers in the anoxic zone, and mechanical surface aerators. Plant B has effluent filtration, but this feature is not required to meet the effluent limits in this design example.

In this example, it is assumed that neither the assimilation of TKN (3.0 mg/L) nor the percent volatile MLSS (63%) will be affected by the solids retention time (θ_c) and that the recycle stream characteristics will remain unchanged. Therefore, the mass balance is unchanged.

Table 8.2
Effluent characteristics and design conditions for Plant B

	mg/L Equivalents	
Characteristic	Primary Effluent	Secondary Effluent
VSS	55	9
TSS	80	15
$CBOD_5$	97	3
TCOD	187	33
SCOD	106	20
TN	29.5	26.5
Alkalinity, as $CaCO_3$	120	—

Design Conditions:
Reactor temperature = 15°C.
Reactor MLSS = 3000 mg/L.
MLVSS = 63%.
Reactor pH range = 7.0–7.6.
Reactor minimum DO = 2.0 mg/L.
Secondary effluent NH_4^+-N = 1.0 mg/L.
Secondary effluent Org-N = 2.0 mg/L.
Secondary effluent NO_3^--N = 5.0 mg/L.
Source: US EPA.

Removal requirements across secondary process:
At average annual loadings (concentrations in mg/L equivalents):

BOD_5 removed = (18,925 m^3/day) (97 − 3 mg/L)/1,000 = 1,779 kg/day (3,922 lb/day
TN removed = (18,925 m^3/day) (29.5 − 8.0 mg/L)/1,000 = 407 kg/day (897 lb/day)
TN removed in waste solids by assimilation = (18,925 m^3/day) (3.0 mg/L)/1,000 = 57 kg/day (125 lb/day)

1. *Sizing of nitrification zone*.

Determine the size of the nitrification reactor.
Calculate maximum nitrifier growth rate, $\hat{\mu}_N$ using the following Arrhenius type expression:

$$\hat{\mu}_N = 0.47\, e^{0.098\ (T-15)}$$

For T = Temperature = 15°C

$$\hat{\mu}_N = 0.47/\text{day}$$

Calculate specific growth rate, μ_N using the Monod kinetic equation:

$$\mu_N = \hat{\mu}_N \frac{N}{K_N + N}$$

For:

N = substrate nitrogen concentration = 1.0 mg/L
K_N = 1.0 mg/L

$$\mu_N = 0.47 \frac{1}{1+1} = 0.23/\text{day}$$

Calculate the minimum solids retention time, θ_c^m

$$\theta_c^m = \frac{1}{\mu_N}$$

$$\theta_c^m = 1/0.23 = 4.35 \text{ day}$$

Calculate design solids retention time, θ_c^d using a design factor

$$\theta_c^d = (\text{PF})\ (\text{SF})\ (\theta_c^m)$$

where:

PF = Peaking Factor = 1.56
SF = Safety Factor = 1.25
$\theta_c^d = (1.56)\ (1.25)\ (4.35) = 8.5$ day

Calculate the organic removal rate, q_{OBS}

$$q_{OBS} = \frac{1}{\theta_c^d Y_{NET}}$$

From Appendix F at θ_c^d of 8.5 day, Y_{NET} is 0.24 g total VSS/g COD removed

$$q_{OBS} = \frac{1}{8.5 \times 0.24} = 0.49 \text{ g COD/g MLVSS/day}$$

Determine reactor hydraulic retention time, t

$$t = \frac{S_o - S_1}{q_{OBS} X}$$

For:

$S_o = 187$ mg/L COD
$S_1 = 20$ mg/L COD
X = MLVSS = 3000 × 0.63 = 1,890 mg/L
$t = (187 - 20)/(0.49 \times 1890) = 0.18$ day = 4.3 h

Calculate reactor volume, V_N:

$$V_N = Q\ t$$

Q = 18,925 m^3/day (5 MGD)
t = 0.18 d = 4.3 h
$V_N = 18{,}925 \times 0.18 = 3410\ m^3$ (0.90 MG)

2. *Determine various design parameters to check validity.*

Actual Retention time, t:

$t = V_{N/Q} = (3{,}410/21{,}056) \times 24 = 3.9$ h at an actual flow of 21,056 m^3/day (5.56 MGD)
$t = 3.9/1.5 = 2.6$ h at $Q_{max\ mo}$ of 31,584 m^3/day (8.34 MGD)

Food/Mass:

$F/M = (Q_{avg} \times BOD_5)/(V_N \times MLVSS)$
$F/M = (18{,}925 \times 97)/(3{,}410 \times 1{,}890) = 0.28\ g\ BOD_5/g\ MLVSS/day$

Specific nitrification rate required, SNR_{min}:

Nitrification required $= 29.5 - 3.0 - 2.0 - 1.0 = 23.5\ mg/L\ NH_4^+$-N

At design average:

$SNR_{min} = (18{,}925 \times 23.5)/(3{,}410 \times 1{,}890) = 0.069\ g\ NH_4^+$-N/g MLVSS/day

At max. monthly:

$SNR_{min} = 0.069 \times PF$, where PF = 1.2 for influent TKN (Appendix G)
$SNR_{min} = 0.069 \times 1.2 = 0.083\ g\ NH_4^+$-N/g MLVSS/day

Check minimum rates required against actual rates measured by testing. If rate tests are not performed, check rates reported in the literature for similar θ_c and COD:TKN or $CBOD_5$:TKN ratios. In this case, $\theta_c = 8.5$ days, COD:TKN = 6.3, and $CBOD_5$:TKN = 3.3. If minimum rates calculated above are too low, increase the process design factor (DF) for sizing reactor.

3. *Sizing of anoxic zone for denitrification.*

Size anoxic zone based on SDNR (specific denitrification rate) and adjust based on design temperature and selected PF for design condition. This example is based on the maximum monthly.

As previously determined, nitrates produced in nitrification zone = 23.5 mg/L equivalents
Denitrification required = 23.5 − 5.0 = 18.5 mg/L or 350 kg NO_3^--N/day
Total NO_3^--N available in recycle (internal and RAS) streams = 23.5 mg/L

Select SDNR from rate tests, rates reported in the literature under similar conditions. In this example, determine the denitrification rate from Fig. 8.1 (40), which shows the specific rate of nitrate removal as a function of the F/M ratio in the anoxic zone, and use it as the average rate for staged compartments. By trial and error:

SDNR = 0.09 g NO_3^--N/g MLVSS/day at 20°C with wastewater as carbon source at anoxic F/M of 0.34 g BOD/g MLVSS/day at maximum monthly conditions.

Adjust nitrate removal rate at $\theta = 1.08$ (1) for T = 15°C:

$SDNR_{T=15} = (0.09)\ (\theta^{T-20}) = (0.09)\ (1.08)^{15-20} = 0.061\ g\ NO_{3^-}$-N/g MLVSS/d

For maximum month where PF = 1.2:

MLVSS required = 350 kg NO_{3^-}-N/day × 1.2PF ÷ 0.061 g NO_{3^-}-N/g MLVSS/day = 6885 kg (15,179 lb)

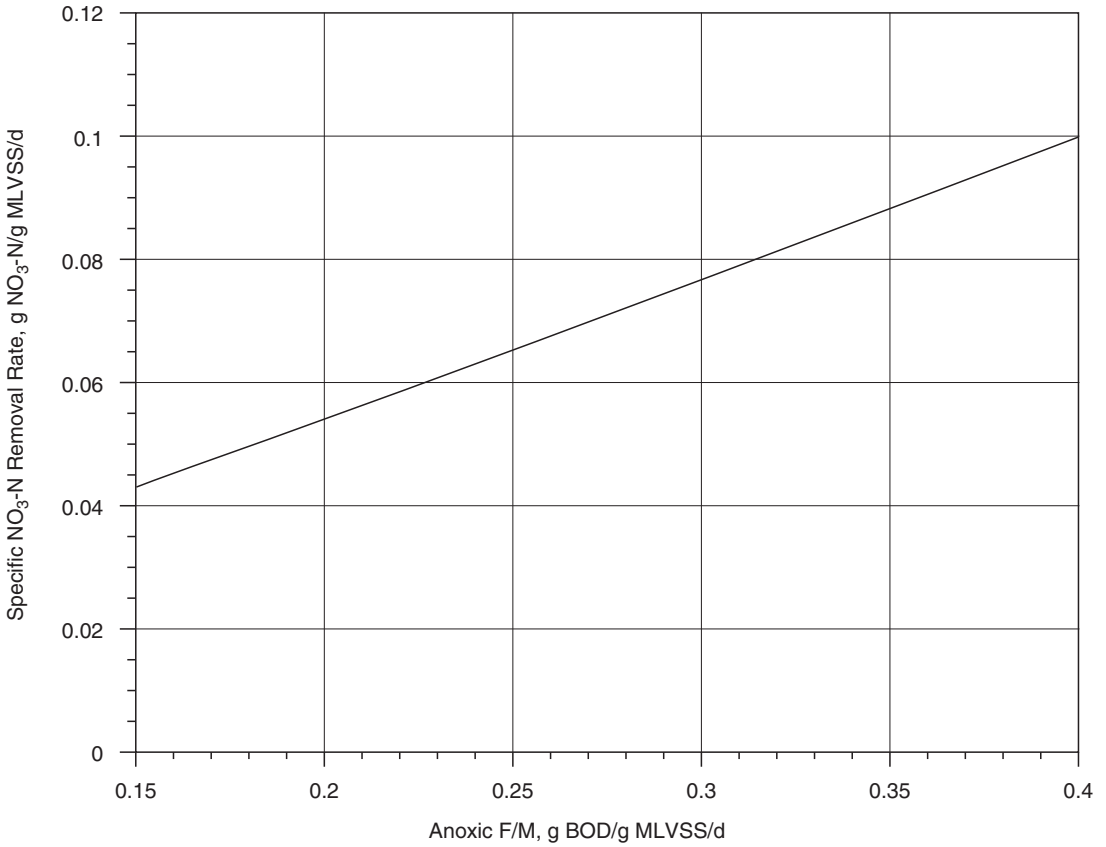


Fig. 8.1. Denitrification rate as a function of Anoxic F/M. (*Source*: US EPA).

Anoxic volume, V_{AN}:

for MLVSS $= 1890\,\mathrm{mg/L}$,

$$V_N = \frac{6885 \times 10^3}{1890} = 3643\,\mathrm{m}^3\ (0.96\,\mathrm{MG})$$

Calculate hydraulic retention time in anoxic zone at actual flow:

At Q_{avg},

$t = (3643/21{,}056) \times 24 = 4.2\,\mathrm{h}$

At $Q_{\mathrm{max\ mo}}$,

$t = 4.2/1.5 = 2.8\,\mathrm{h}$

System θ_c^d = aerobic θ_c^d + anoxic $\theta_c^d = 8.5 + \dfrac{3643}{3410}(8.5) = 17.6$ day

4. *Determine RAS rate.*

To maintain MLSS = 3000 mg/L at Q_{RAS} = 7000 mg/L

$$Q_{RAS} = \frac{3000Q}{(7000 - 3000)} = 0.75Q$$

At Q = 21,056 m^3/day (5.563 MGD),
Q_{RAS} = 15,790 m^3/day (4.17 MGD)
At $Q_{max\ mo}$ = 31,584 m^3/day (8.34 MGD),
Q_{RAS} = 23,690 m^3/day (6.26 MGD)

5. *Size anoxic internal recycle rate Q_1.*

From previous calculations of denitrification required and NO_3^--N returned to anoxic zone, the denitrification efficiency required equals 18.5/23.5 × 100, which is 78.7%

For single anoxic zone system:

$$\text{Denitrification efficiency} = \frac{Q_1 + Q_{RAS}}{Q + Q_1 + Q_{RAS}} \times 100\%$$

$$0.787 = \frac{Q_1 + 0.75Q}{Q + Q_1 + 0.75Q}$$

$Q_1 = 2.95Q$
At $Q_{max\ mo}$ = 31,584 m^3/day (8.34 MGD),
Q_1 = 93,160 m^3/day (24.6 MGD)

6. *Determine alkalinity requirements to maintain residual alkalinity of 50 mg/L equivalents as $CaCO_3$ with influent alkalinity = 120 mg/L as $CaCO_3$.*

Alkalinity demand = (7.14 mg $CaCO_3$/mg NH_4^+-N) (23.5 mg/L NH_4^+-N oxidized)
= 168 mg/L as $CaCO_3$
Supplemental alkalinity addition required = (168 + 50) − 120 = 98 mg/L as $CaCO_3$
Savings with denitrification = (3.6 mg $CaCO_3$/mg NH_4^+-N reduced) (18.5) = 67 mg/L as $CaCO_3$
Average supplemental alkalinity required with denitrification = 98 − 67 = 31 mg/L as $CaCO_3$

Size the maximum capacity of the feed system on peak conditions in a similar fashion to prevent violation of pH limits, usually daily. Size the system for peak day demand.

7. *Determine mixing requirements in anoxic zone.*

At 50 hp/MG of anoxic volume, the minimum total hp required equals 48 where V_{AN} = 3,643 m^3 (0.96 MG).

The number and size of each mixer is based on the number of anoxic compartments and the compartment's configuration: Verify mixing requirements based on reactor depth and configuration. With six compartments, 8 hp is required per compartment. Therefore, each mixer is 10 hp for next standard size unit.

8. *Determine waste sludge requirements* (as outlined in US EPA Manual Ref. 1).

$$\theta_c^d = \frac{l_A}{S}$$

where

l_A = VSS under aeration, and
S = VSS wasted daily
$l_A = (1{,}890\,\text{mg/L MLVSS} \times 3{,}410\,\text{m}^3)/1{,}000 = 6{,}445\,\text{kg MLVSS}$ (14,210 lb MLVSS)
$S = 6{,}445/8.5 = 758\,\text{kg VSS/day}$ (1,670 lb VSS/day) to be wasted
Volatile solids contained in the effluent = 9 mg/L (from Table 8.2)
Sludge wasting from RAS $= 758 - (9)(18{,}925)/10^3 = 607\,\text{kg VSS/day}$ (1,340 lb VSS/day)

At MLVSS/MLSS = 63%:

Average WAS = 963 kg TSS/day (2,120 lb/day TSS)
Determine WAS pumping rate at X_{RAS} = 7000 mg/L:

$$\text{Average } Q_{WAS} = \frac{963(10^3)}{7{,}000} 138\,\text{m}^3/\text{day}\ (36{,}400\,\text{gpd})$$

Similarly, both WAS mass and WAS pumping requirements should be determined from mass balances at peak conditions.

9. *Determine aeration requirements under various design conditions.*

Average Conditions

BOD_5 removed $= 18{,}925\,\text{m}^3/\text{day}\ (97–3)\,\text{mg/L}/10^3 = 1779\,\text{kg/day}$ (3922 lb/day)
NH_4^+-N oxidized $= 18{,}925\,\text{m}^3/\text{day} \times 23.5\,\text{mg/L}/10^3 = 445\,\text{kg/day}$ (981 lb/day)

For this example, 1.1 kg O_2/kg BOD_5 removal was assumed for the carbonaceous demand. The range for nitrification systems is about 1.0 to 1.3, depending on θ_c and the temperature.

Total oxygen demand $= (1.1 \times 1779) + (4.6 \times 445) = 4004\,\text{kg/day}$ (8,827 lb/day)

Peak day conditions (peaking factors from Appendix G):

Total oxygen demand
$= (2.1 \times 1.1 \times 1779) + (1.7 \times 4.6 \times 445) = 7589\,\text{kg/day}$ (16,730 lb/day)

Savings in O_2 demand with denitrification from wastewater BOD_5:

O_2 saved $= 2.9\,\text{mg } O_2/\text{mg } NO_3^- - N \times 18.5\,\text{mg/L } NO_3^- - N = 53.7\,\text{mg/L}$ or 1015 kg/day (2237 lb/day) avg

Oxygen demand with denitrification:

Avg. O_2 demand $= 4004 - 1015 = 2989\,\text{kg/day}$ (6590 lb/day)
Peak day O_2 demand $= 7589 - (1.7)(1015) = 5864\,\text{kg/day}$ (12,926 lb/day)

Mechanical aeration sizing:

At 50 lb O_2/hp/day for mechanical aerators, the minimum hp required is 259. For series arrangement of aeration basins, aerators should be sized to meet the higher oxygen demand in the first basin. Two-speed motors and adjustable submergence are recommended to meet the varying conditions.

10. *A typical flow configuration with baffles in the anoxic zone is shown in Fig. 8.2.*

Nitrified flow is recycled from the end of the nitrification zone to the head of the first anoxic compartment. Six mixers required at 10 hp for anoxic compartments. Table 8.3 summarizes reactor volumes at 10°C and 20°C for comparison. The volume changes reflect the impact of temperature, as well as the impact of θ_c on the observed yield, Y_{NET}, from Appendix F.

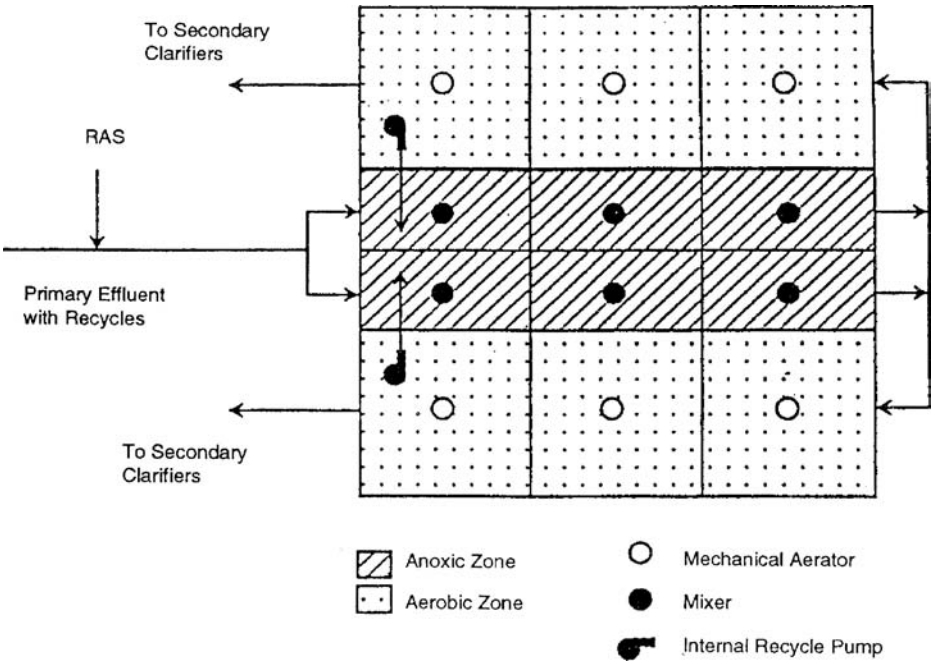


Fig. 8.2. Single anoxic zone system - design example 1. (*Source*: US EPA).

Table 8.3
Volume of reactors at various temperatures—Design Example 1

	Design Temperature		
Volume, m^3	10°C	15°C	20°C
V_{AN}	5370	3640	2270
V_{Nit}	4980	3410	2380
V_{Tot}	10,350	7050	4650

Source: US EPA.

5.3. Design Example 2: Plant B with more Stringent Limits

The inlet wastewater characteristics (feed to secondary system) and secondary effluent characteristics from the mass balance in Appendix D that have been shown in Table 8.1 for Example 1 are the same for the purpose of this example. To meet the more stringent effluent limitations for SS, effluent filtration is required. However, the design conditions are as shown below:

Design Conditions:

Reactor temperature = 15°C	Reactor minimum DO = 2.0 mg/L
Reactor MLSS = 3,000 mg/L	Secondary effluent NH_4^+–N = 1.0 mg/L
Volatile MLSS = 63%	Secondary effluent Org–N = 2.0 mg/L
Reactor pH range = 7.0–7.6	Secondary effluent NO_3^-–N = 1.5 mg/L

To meet the year-round limit of 5 mg/L TN, design for 4.5 mg/L equivalents TN or less at maximum monthly loading and before filtration. Secondary effluent Org-N of approximately 2.0 mg/L equivalents represents the nonbiodegradable fraction of soluble TKN and nitrogen associated with effluent VSS. Approximately 1 mg/L of nitrogen associated with the VSS will be removed by effluent filtration.

With an influent raw TKN concentration of 30 mg/L and an effluent limit of 5 mg/L TN, essentially complete nitrification is required and a very high level of denitrification is required (>90%). To meet this limit, a dual anoxic zone type process configuration is required to provide the additional denitrification required in a second anoxic zone. The plant configuration is as shown in Appendix C with effluent filtration. The process configuration would consist of two anoxic zones and two aerobic zones for nitrogen removal with internal recycle, mixers in the anoxic zones, and diffused aeration.

For this example, it is assumed that the assimilation of NH_4^+-N (3.0 mg/L equivalents) and percent volatile MLSS (63%) will not be affected by the θ_c to remain consistent with the mass balance in Appendix D.

Removal requirements across secondary process at average annual loadings (concentrations in mg/L equivalents):

BOD_5 removed = 1779 kg/d (3922 lb/day)
TN removed by nitrification-denitrification = 473 − 57 = 416 kg/day (917 lb/day)
TN removed in waste solids by assimilation = 57 kg/day (125 lb/day)
TN removed by nitrification-denitrification = 473 − 57 = 416 kg/day (917 lb/day)

1. *Sizing of nitrification zone (first aerobic zone.)*

Following the same procedure used in Example 1, the reactor volume $V_N = 3{,}410\,m^3$ (0.90 MG) using the same design factor. Because the limits are more stringent and year round, a more conservative design factor may be considered if daily or seasonal variations are significant.

2. *Sizing of first anoxic zone for denitrification.*

Size first anoxic zone based on specific denitrification rate with wastewater as the carbon source and adjust based on design temperature and selected PF.

Allowing 1.5 mg/L equivalents of NO_3^--N in the final effluent, the denitrification required with both anoxic zones = 23.5 − 1.5 = 22.0 mg/L or 416 kg NO_3^--N/day (917 lb NO_3^--N/day).

Nitrates produced in first aerobic zone = 23.5 mg/L equivalents
NO_3^--N in internal recycle stream to first anoxic zone = 23.5 mg/L equivalents

Determine the maximum percent denitrification removal in the first anoxic zone based on the practical limit for internal recycle from the end of the first aerobic zone to the first anoxic zone.

Select $Q_1 = 450\%$ or $4.5Q$ as practical limit. Although Q_1 can be higher, denitrification efficiency increases at a decreasing rate (as shown in Fig. 8.8, Chapter 7).

For dual anoxic zone systems, any return of nitrates to the anoxic zone from Q_{RAS} will be small because of the removals in the second anoxic zone and can be ignored.

Denitrification efficiency = $\frac{Q_1}{Q+Q_1+Q_{RAS}}(100\%)$
$Q_{RAS} = 0.75\ Q$ (from Example 1)
Denitrification efficiency = $\frac{4.5Q}{Q+4.5Q+0.75Q}(100\%) = 72\%$ maximum
NO_3^-−N removed in first anoxic zone = 0.72 × 23.5 mg/L = 16.9 mg/L or 320 kg/day (706 lb/day)
NO_3^-−N removal required in second anoxic zone = 5.1 mg/L or 96 kg/day (213 lb/day)

Select SDNR as in Example 1. Because the amount of denitrification required in the first anoxic zone is similar to that in Example 1, the anoxic F/M and SDNR will be approximately the same.

At 15°C, $SDNR_1 = (0.09)\ (1.08)^{-5} = 0.061$ g NO_3^-−N/g MLVSS/day
For maximum month, MLVSS required = 320 kg NO_3^- − N/day × 1.2 PF/0.061 g NO_3^- − N/g MLVSS/day
= 6295 kg (13,880 lb)

$V_{AN1} = 6295$ kg MLVSS × 10^3/1890 mg/L MLVSS = 3331 m^3 (0.88 MG)

Calculate actual retention time

At Q of 21,056 m^3/day (5.56 MGD), $t = V_{AN1}/Q = (3.331\ m^3/21{,}056\ m^3/day) \times 24 = 3.8$ h
At $Q_{max\ mo}$ of 31,584 m^3/day (8.34 MGD), $t = (3.331\ m^3/31{,}584\ m^3/day) \times 24 = 2.5$ h

3. Determine RAS rate.

$Q_{RAS} = 0.75\ Q$ similar to Example 1

4. Internal recycle rate.

As determined previously, $Q_1 = 4.5\ Q$

At maximum monthly flow, $Q_1 = 142{,}130\,m^3/day$ (37.5 MGD)

5. *Size second anoxic zone for denitrification with endogenous carbon.*

Determine $SDNR_2$ from the following equation as a function of θ_c (1):

$SDNR_2 = 0.12\ \theta_c^{-0.706}$ at 20°C

By trial and error, estimate that the system $\theta_c = 25$ days.

$SDNR_2 = 0.12 \times 0.103 = 0.0124\,g\ NO_3^- - N/g$ MLVSS/day

Adjust for $T = 15°C$ using Arrhenius temperature constant = 1.03

$SDNR_2 = 0.0124 \times (1.03)^{15-20} = 0.011\,g\ NO_3^- - N/g$ MLVSS/day

$NO_3^- - N$ removal required in second zone = 96 kg/day (213 lb/day), as previously determined.

MLVSS required = (96 kg/day)(96 kg/day $NO_3^- - N$)/(0.11 g $NO_3^- - N$/g MLVSS/day = 8,727 kg (19,240 lb)

$V_{AN2} = (8727\,kg\ MLVSS \times 1.2 \times 10^3)/1890\,mg/L\ MLVSS = 5541\,m^3$ (1.47 MG)

At Q of $21{,}056\,m^3$/day (5.563 MGD), $t = V_{AN2}/Q = 6.3$ h

At $Q_{max\ mo}$, $t = 4.2$ h

6. *Determine alkalinity requirements as in Example 1.*
7. *Determine mixing requirements as in Example 1.*

50 hp/MG required for mixing in the anoxic zones

$V_{AN1} + V_{AN2} = 3331 + 5541 = 8872\,m^3$ (2.35 MG)

Total hp required = $50 \times 2.35 = 118$ hp

8. *Determine waste sludge as in Example No. 1.*
9. *Size postaeration zone.*

Size for 0.5 h detention time at $Q_{max\ mo} = 31{,}584\,m^3$/day (8.34 MGD).

$V_{PA} = 660\,m^3$ (0.17 MG)

10. Calculate aeration requirements with fine-pore diffused aeration system.

Similar to Example 1 for carbonaceous oxidation and nitrification, but with postaeration zone to raise DO from 0 to 2.0 mg/L.

For average day, total oxygen demand = 4004 + 42 = 4046 kg/day (8920 lb/day)

For peak day, total oxygen demand = 7684 kg/day (16,940 lb/day)

Savings in O_2 demand with denitrification:

O_2 savings = (2.9 mg O_2/mg $NO_3^- - N$ denit.) (22 mg/L $NO_3^- - N$ denit.) = 63.8 mg/L or 1207 kg/day (2660 lb/day) avg.

Oxygen demand with denitrification:

For average day, total oxygen demand = 4046 − 1207 = 2839 kg/day (6260 lb/day)

For peak day, total oxygen demand = 7684 − (1.7)(1207) = 5632 kg/day (12,416 lb/day)

At 12.5% O_2 transfer efficiency—assumed at 4.6 m (15 ft) diffuser submergence:

Peak air required = (5632 kg O_2/day)/[(0.125)(0.28 kg O_2/m^3 air)(1440 min/day)]
= 112 m^3/min (3955 cfm)
Provide three blowers plus one standby
Blower capacity = 38 m^3/min/blower (1340 cfm/blower)

Check mixing requirements at end of nitrification zone with tapered aeration system to ensure that the air provided for oxygen demand is adequate to meet the mixing requirement.

11. Reactor configuration

Reactor configuration is similar to reactor configuration in Example 1, but add second anoxic zone and post aeration as shown in Fig. 8.3. Nitrified flow is recycled from the end of the nitrification zone to the head of the first anoxic compartment. Table 8.4 summarizes reactor volume requirements at 10°C, 15°C, and 20°C for comparison:

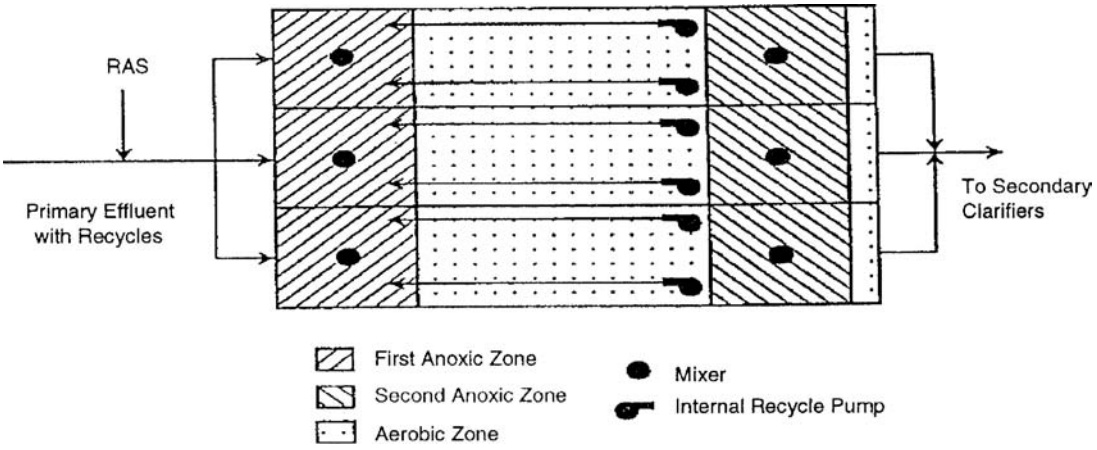


Fig. 8.3. Dual anoxic zone system - design example 2. (*Source*: US EPA).

Table 8.4
Volume of reactors at various temperatures—Design Example 2

	Design Temperature		
Volume, m^3	10°C	15°C	20°C
V_{AN1}	4950	3330	2110
V_{Nit}	4980	3410	2380
V_{AN2}	6450	5540	4670
V_{PA}	660	660	660
V_{Tot}	17,040	12,940	9820

Source: US EPA.

With diffused aeration, use a tapered aeration pattern to match higher O_2 demand at the head end of the nitrification zone. Provide one mixer for each anoxic compartment. For six compartments, each mixer would be 20 hp.

12. To meet the TP limitation of 1.0 mg/L for this process configuration

The logical process selection would be to incorporate biological phosphorus removal with nitrogen removal followed by chemical addition for phosphorus polishing to meet the limit consistently. This process can be accomplished by adding an anaerobic selector ahead of the first anoxic zone, typically 1 to 2 hours nominal retention time. The internal recycle of nitrified flow would continue to be returned to the first anoxic zone. RAS, however, would be recycled to the head of the anaerobic selector. To operate at maximum efficiency for biological phosphorus removal, it would be necessary to operate at the minimum θ_c necessary to achieve nitrification and at maximum denitrification efficiency to minimize the return of nitrates to the anaerobic selector. In addition, the internal recycle rate and anoxic zone volume in use must be carefully monitored to prevent an excessive anoxic period (i.e., inadequate $N0_3{}^-$-N to denitrify) as this can cause an excess release of phosphates—a secondary release phenomenon, which occurs without storage of BOD.

Consequently, extensive process monitoring and process control are required to maintain proper recycle rates, solids inventory, DO control, and sludge blanket levels. Polishing would be accomplished by adding chemicals such as ferric chloride and alum to the postaeration zone ahead of the secondary clarifiers followed by effluent filtration for SS removal to meet the limit of 1 mg/L TP.

To meet the phosphorus removal requirement by chemical precipitation only, the impact on the nitrogen removal system would primarily result from an increase in sludge production (chemical sludge) and reduced fraction of volatile solids in the mixed liquor. As the system would be limited in its ability to carry MLSS, the volume of aerobic and anoxic tankage could increase significantly, as well as pumping requirements for WAS and RAS.

The use of metal salts to precipitate phosphorus will cause a loss of alkalinity. Therefore, an increase in supplemental alkalinity addition could be required to maintain a residual alkalinity of 50 mg/L as $CaCO_3$.

5.4. *Design Example 3—Plant A with Less Stringent Limits*

From the mass balances (Appendix E) the inlet wastewater characteristics for the less complex plant (raw influent plus recycles) and secondary effluent characteristics without nitrification-denitrification are as shown in Table 8.5.

To meet seasonal limits of 2 mg/L $NH4^+$-N and 10 mg/L TN at maximum monthly loading, design for 1.0 mg/L equivalents $NH_4{}^+$-N and 8.0 mg/L equivalents TN. Secondary effluent Org-N of approximately 2.0 mg/L equivalents represents the nonbiodegradable fraction of soluble TKN and nitrogen associated with effluent VSS.

With raw influent TKN = 30 mg/L and effluent limits of 2 mg/L $NH_4{}^+$-N and 10 mg/L TN, essentially complete nitrification and a minimum of 67% TN removal efficiency are required. To meet these limits, a single anoxic zone or phased system, such as the cyclical aeration process, can be used. The plant configuration for the less complex plant A will use cyclical

Table 8.5
Effluent characteristics and design conditions for Plant A

	mg/L Equivalents	
Characteristic	Aeration Tank Feed	Secondary Effluent
VSS	129	9
TSS	187	15
$CBOD_5$	152	3
TCOD	290	33
SCOD	100	20
TN	30.5	25.2
Alkalinity, as $CaCO_3$	120	–

Design Conditions:
Reactor temperature = 15°C.
Reactor MLSS = 3000 mg/L.
MLVSS = 58%.
Reactor pH range = 7.0–7.6.
Aerobic phase DO = 2.0 mg/L.
Secondary effluent NH_4^+-N = 1.0 mg/L.
Secondary effluent Org-N = 2.0 mg/L.
Secondary effluent NO_3^--N = 5.0 mg/L.
Source: US EPA.

nitrogen removal with mechanical aerators to provide alternating aerobic anoxic cycles within the same basin with mixing during the off-cycle.

Removal requirements across secondary process:

At average annual loadings (concentrations in mg/L equivalents)
BOD_5 remaining = (18,925 × 149)/1,000 = 2,820 kg/d (6,217 lb/day)
TN remaining = [18,925 (30.5 − 8)]/1,000 = 426 kg/d (939 lb/day)
TN remaining in waste solids by assimilation = $18{,}925 \times 5.3/10^3 = 100$ kg/d (221 lb/day)
TN remaining required by nitrification-denitrification 426 − 100 = 326 kg/day (719 lb/day) or 17.2 mg/L
NH_4^+−N produced = 30.5 − 5.3 − 2.0 − 1.0 = 22.2 mg/L
Denitrification required during anoxic phase = 17.2 mg/L

1. *Sizing of nitrification or aerobic phase*.

Similar to Example 1, $\theta_c^d = 8.5$ days for nitrification. Without separate digestion facilities, however, assume that sludge stabilization is required within the aeration tanks. For this example, to achieve stabilization, choose $\theta_c^d = 20$ days for 10°C to 20°C. Following the same procedure as in Example 1, the following calculation can be made:

$$Q_{OBS} = 1/(20 \times 0.27) = 0.185 \text{ g COD/g MLVSS/day}$$

$$t = \frac{S_0 - S_l}{q_{OBS} X} = \frac{290 - 20}{0.185 \times 3000 \times 0.58} = 0.84 \text{ days}$$

$V_{aer} = 0.84 \times 18{,}925 = 15{,}900\,m^3$ (4.20 MG); or $6{,}760\,m^3$ (1.78 MG) required for nitrification

where $Q = 18{,}925\,m^3$/day was used to determine required aeration volume because t was computed based upon milligram per liter equivalent concentrations.

2. *Check design parameters.*

At $Q = 19{,}379\,m^3$/day (5.12 MGD), the actual retention time $t = 19.7$ h

$$F/M = \frac{18{,}925 \times 152}{15{,}900 \times 1740} = 0.10\ g\ BOD/g\ MLVSS/day$$

3. *Sizing of anoxic phase for denitrification.*

Size based on specific denitrification rate, and adjust based on design temperature and selected PF. For a cyclical system, the denitrification rate can vary between the rate in a dedicated anoxic basin with internal recycle and with wastewater as the carbon source, and the rate with endogenous decay. At the beginning of the off-cycle, the unmetabolized wastewater COD level will be very low. At the midpoint of the off-cycle, where the off-cycle duration equals one-half the retention time in that basin, the COD level will be similar to a dedicated anoxic zone with recycle. With the high COD:TKN for this example, it is estimated that the overall SDNR will equal the endogenous rate for 25% of the anoxic phase and the wastewater carbon rate for 75% of the cycle.

Because the quantity of denitrification required is approximately the same as in the first anoxic zone of Example 2 but the rate is less, assume the anoxic F/M will be similar. Therefore, determine SDNR from weighted average. From the previous example, $SDNR_2 = 0.011$ g NO_3^--N/g MLVSS/day and $SDNR_1 = 0.061$ g NO_3^--N/g MLVSS/day. The weighed average is 0.049 g NO_3^--N/g MLVSS/day.

Denitrification required = 326 kg/day (719 lb/day)
For maximum month, MLVSS required = (326 × 1.2 PF)/0.049 = 7,984 kg (17,601 lb)
$V_{AN} = (7984 \times 103)/1740 = 4589\,m^3$(1.21 MG)
Check anoxic F/M:
F/M = (18,925)(152)/(4,589)(1740)] = 0.36, which is approximately equal to assumed 0.34 from Example 1 for trial-and-error solution.
Calculate actual hydraulic retention time for anoxic phase:
At Q = 19,379 m^3/day (5.12 MGD), t = 5.7 h
At $Q_{max\ mo}$, $t = 3.8$ h

4. *Determine ratio of anoxic and nitrification periods (cycling of on/off periods).*

$V_{Nit} = 6760\,m^3$ (1.78 MG)
$V_{AN} = 4{,}589\,m^3$ (1.21 MG)
On/off ratio=3:2

With two trains and three cycled compartments and one final continuously aerated compartment per train, the retention time in each cycled compartment is approximately 4.69 h. Off-cycle duration should equal at least one-half the retention time in each compartment to maintain high average COD:TKN during the anoxic cycle. The basins following the cycled basins would be aerated continuously for sludge stabilization.

Off-cycle = 2.4 h
On-cycle = 3.6 h

5. *Determine RAS rate similar to previous example.*

No internal recycle is required if step-feeding to downstream compartments is provided to allow use of wastewater carbon as the primary carbon source for denitrification.

$$Q_{RAS} = 0.75Q$$

6. *Determine alkalinity requirements as in Example 1*
7. *Determine mixing requirements during anoxic phase*

Mixers are desirable to obtain full liquids-solids contact while the air is off and to distribute raw wastewater carbon source for denitrification particularly for downstream compartments with step feed.

8. *Calculate waste sludge and aeration requirements similar to previous examples*

The aeration rate required for each cycled basin, however, should be increased by 67% to account for the off-cycle. The aeration required for the final basins would be based on sludge stabilization requirements.

9. *Flow configuration*

The flow configuration is shown in Fig. 8.4. The first three basins in each train would have cycled aeration and the final basin would be continuously aerated. Step feeding of influent is provided to each of the three cycled basins. Table 8.6 summarizes reactor volumes at 10°C, 15°C, and 20°C for comparisons.

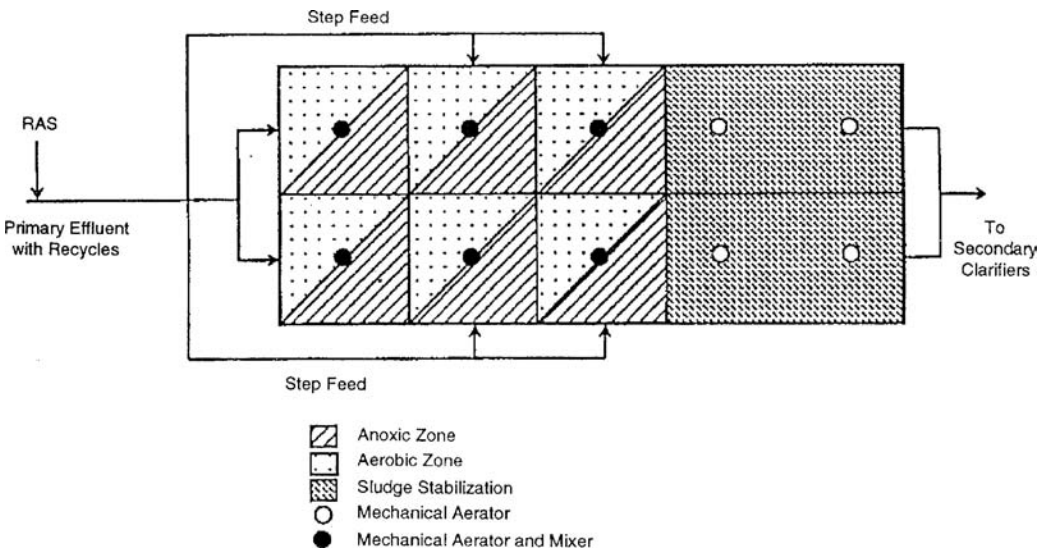


Fig. 8.4. CNR process - design example 3. (*Source*: US EPA).

Table 8.6
Volume of reactors at various temperatures—Design Example 3

	Design Temperature		
Volume, m^3	10°C	15°C	20°C
V_{Nit}	9870	6760	4250
V_{AN}	6400	4590	3100
V_{PA}	6020	9140	11640
V_{Tot}	22,290	20,490	18,990

Source: US EPA.

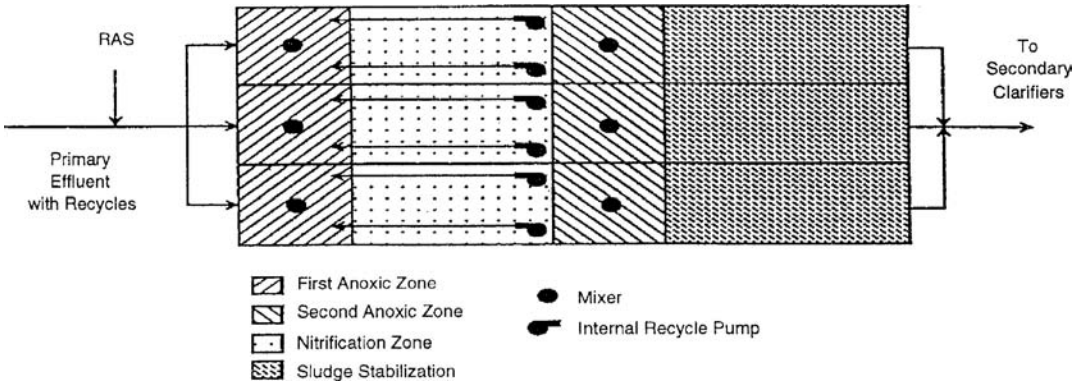


Fig. 8.5. Dual anoxic zone system - design example 4. (*Source*: US EPA).

The total volume of each cycled compartment is the sum of V_{Nit} and V_{AN} and the total volume required for sludge stabilization is the sum of V_{Nit} and V_{PA}. The total volume for sludge stabilization should be based on state and federal guidelines or requirements for stabilization at various temperatures.

5.5. Design Example 4—Plant A with More Stringent Limits

To meet the more stringent limitations with the less complex plant configuration, again a dual anoxic zone type of process configuration would be required similar to that in Example 2 as shown in Fig. 8.5. The procedures for sizing would be identical to those used in Example 2 except the total aerobic volume would be governed by the volume required to achieve sludge stabilization. The volume required is summarized for various temperatures in Table 8.7.

NOMENCLATURE

DF = process design factor
DO = dissolved oxygen (concentration or weight), mg/L or kg
F/M = food to microorganisms ratio, g BOD_5/g MLVSS/day (lb BOD_5/lb MLVSS/day)

Table 8.7
Volume of reactors at various temperatures—Design Example 4

	Design Temperature		
Volume, m^3	10°C	15°C	20°C
V_{AN1}	5040	3390	2150
V_{Aer}	9870	6760	4250
V_{AN2}	8370	7190	6060
V_{PA}	6020	9140	11,640
V_{Tot}	29,300	26,480	24,100

Source: US EPA.

K_N = half reaction rate constant for nitrification, mg/L
l_A = VSS under aeration, kg (lb)
MLSS = mixed liquor suspended solids (concentration or weight) mg/L or kg
MLVSS = mixed liquor volatile suspended solids (concentration or weight), mg/L or kg
N = substrate nitrogen concentration, mg/L
PF = Peaking Factor
q_{OBS} = organic removal rate, g COD/g MLVSS/day
Q = flow rate, m^3/day (MGD or gpm)
Q_1 = anoxic internal recycle rate, m^3/day (MGD or gpm)
Q_{avg} = average floe rate, m^3/day (MGD or gpm)
$Q_{max\ mo}$ = maximum monthly flow rate, m^3/day (MGD or gpm)
Q_{OBS} = observed flow rate, m^3/day (MGD or gpm)
Q_{RAS} = return activated sludge rate, m^3/day (MGD or gpm)
Q_{WAS} = wasted activated sludge rate, m^3/day (MGD or gpm)
S = VSS wasted daily, kg VSS/day (lb VSS/day)
SDNR = specific denitrification rate, g NO_3^--N/g MLVSS/day
SF = Safety factor
SNR = specific nitrification rate, g NH_4^+-N/g MLVSS/day
SDNR = specific denitrification rate, g NO_3^-/g MLVSS/day
SNR_{min} = minimum specific nitrification rate, g NH_4^+-N/g MLVSS/day
t = reactor hydraulic retention time, h
T = reactor temperature, °C (°F)
V_{aer} = aerobic reactor volume, m^3 (MG)
V_{AN} = anoxic reactor volume, m^3 (MG)
V_{PA} = size post-aeration zone, m^3 (MG)
$V_N = V_{Nit}$ = nitrification reactor volume, m^3 (MG)
V_{Tot} = Total reactor volume
X = MLVSS, mg/L
X_{RAS} = return activated sludge concentration, mg/L
Y_{NET} = Yield, VSS/g COD removed

μ_N = specific growth rate, 1/day
$\hat{\mu}_N$ = maximum nitrifier growth rate, 1/day
θ_c = solids retention time, day
θ_c^m = minimum solids retention time, day
θ_c^d = design solids retention time, day

REFERENCES

1. US EPA, *Nitrogen Control Manual*, US Environmental Protection Agency, Tech. Report # EPA/625/R-93/010, Office of Research and Development, Washington, DC, September (1993).
2. N. K. Shammas, Interactions of temperature, pH and biomass on the nitrification process, *Journal Water Pollution Control Federation*, 58, 1, 52–59, January (1986).
3. N. K. Shammas, An allosteric kinetic model for the nitrification process, *Proceedimgs of the Tenth Annual Conference of Water Supply Improvement Association*, Honolulu, Hawaii, pp. 1–30, July (1982).
4. N. K. Shammas, *Optimization of Biological Nitrification, Ph.D. dissertation*, Microfilm Publication, University of Michigan, Ann Arbor, Michigan (1971).
5. L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), *Biological Treatment Processes*, Humana Press, Totowa, NJ (2009).
6. C. Beer, and L. K. Wang, Activated sludge systems using nitrate respiration - Design Considerations, *Journal of Water Pollution Control Federation*, 50, 2120, September (1978).
7. C. W. Randall, J. L. Barnard, and H. D. Stensel, Design and retrofit of wastewater treatment plants for biological nutrient removal, Vol. 5, *Water Quality Management Library*, Technomic Publishing, Lancaster, PA (1992).
8. G. T. Daigger, G. D. Waltrip, E. D. Romm, and L. M. Morales, Enhanced secondary treatment incorporating biological nutrient removal, *Journal of Water Pollution Control Federation*, 60, 10, 1833–1842 (1988).
9. C. W. Randall, et al., Retrofitting activated sludge plants to improve the removal of nitrogen and phosphorus by biological process, *Water Science and Technology* (G.B.), 22, 275 (1990).
10. K. Krauth, Practical experiences with nitrification and denitrification processes, *Proceedings of the Japanese-German Workshop on Wastewater and Sludge Treatment*, Tsukuba, October (1982).
11. M. Kuribayashi, Study of nitrified liquor recycled biological nitrification-denitrification processes, *Proceedings of the Japanese-German Workshop on Wastewater and Sludge Treatment*, Tsukuba, October (1982).
12. V. Miyaji, et al., Biological nitrogen removal by step feed process, *Prog. Water Tech*., 12, 6, 193–202 (1980).
13. L. K. Wang, L. Kurylko, and M. H. S. Wang, *Sequencing Batch Liquid Treatment*, US Patent No. 5354458. US Patent & Trademark Office, Washington, D.C (1996).
14. D. S. Lee, C. O. Jeon, J. M. Park, Biological nitrogen removal with enhanced phosphate uptake in a sequencing batch reactor using single sludge system. *Water Res.* 35, 16, 3968–76 November (2001).
15. R. P. G. Bowker, J. M. Smith, and H. Shah, *Assessment of cost and effectiveness of biological dual nutrient removal technologies in the Chesapeake Bay drainage basin*, Vol. 1, EPA CBP/TRS 17, Annapolis, MD (1988).

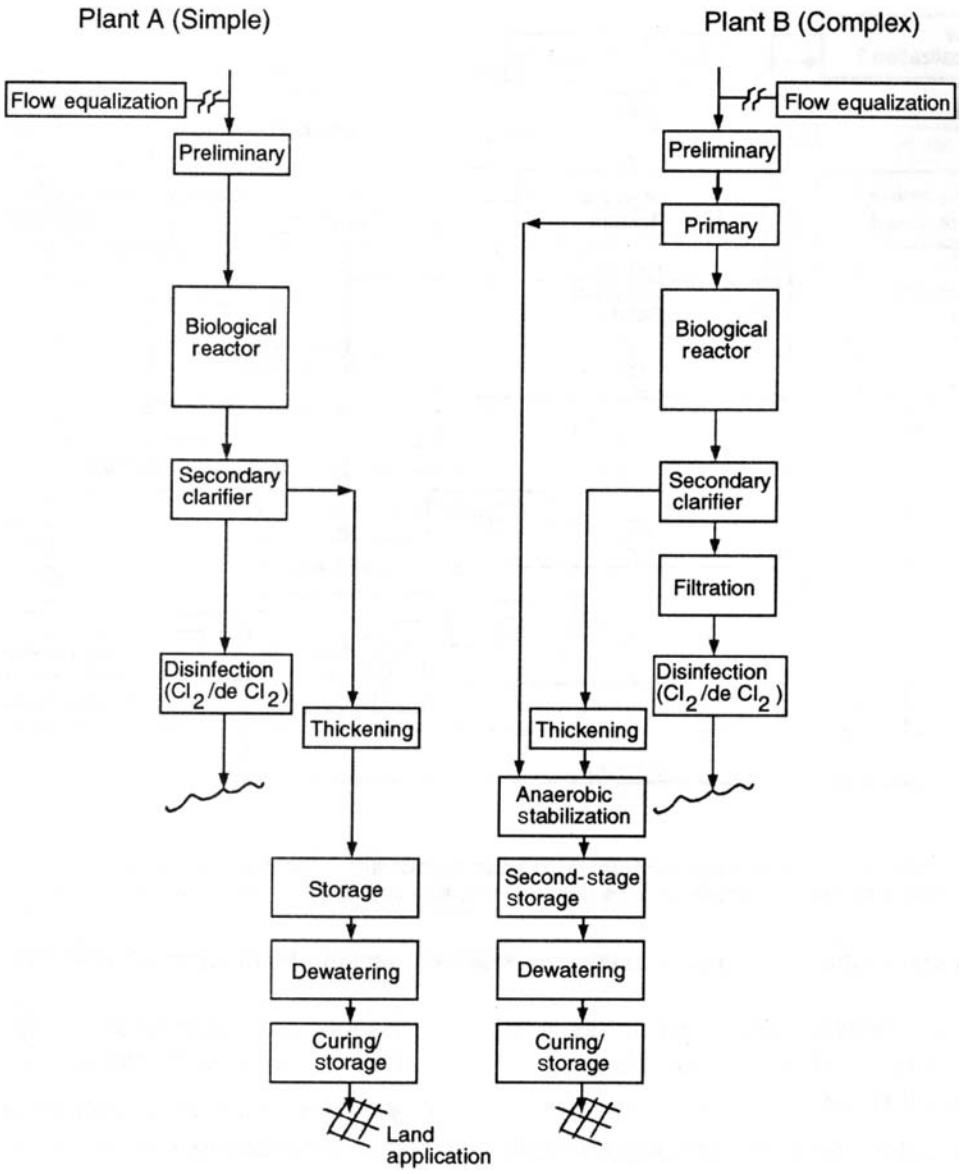
16. J. C. Wang, D. B. Aulenbach, and L. K. Wang, Energy models and cost models for water pollution controls—Chapter 29, *Clean Production*, Misra, K. B. (ed.) Springer-Verlag, Berlin, Germany, pp. 685–720 (1996).
17. L. K. Wang, and N. C. Pereira, *Handbook of Environmental Engineering, Vol. 3, Biological Treatment Processes*, Humana Press, Totowa, NJ, pp. 520 (1986).
18. Soap and Detergent Association, *Principles and Practice of Phosphorus and Nitrogen Removal from Municipal Wastewater*, New York (1991).
19. G. T. Daigger, *Biological Nutrient Removal*, CH2M-Hill, Denver, CO, Paper located on the Web http://bridge.ecn.purdue.edu/~alleman/w3-class/456/article/daigger-nutrients/daigger.html, (2005).
20. R. O. Jr. Mines, *Designing Single-Sludge Bionutrient Removal Systems*, 2001 World Water and Environmental Resources Conference, Orlando, FL (2001).
21. M. C. Goronszy, CASS, *Cyclic Activated Sludge System: Superior Batch Reactor Technology*, Transenviro, Inc., Aliso Viejo, CA (1990).
22. IAWPRC, *Simulation of Single-Sludge Processes for Carbon Oxidation, Nitrification and Denitrification*, Model developed by the IAWPRC Task Group on Mathematical Modeling for Design and Operation of Biological Wastewater Treatment, Version 1, 1997, Clemson University, Source Link: http://www.ces.clemson.edu/ees/sssp January (2004).
23. Y. Argaman, G. Papkov, A. Ostfeld, and D. Rubin, Single-sludge nitrogen removal model: calibration and verification, *ASCE, J. Envir. Engrg*., 125, 7, 608–617, July (1999).
24. C. G. Jih, J. S. Huang, and K. C. Hsieh, Performance evaluation of single-sludge reactor system treating high-strength nitrogen wastewater, *Journal of Hazardous Materials*, 85, 3, 213–27, August (2001).
25. BBS, *Single-Sludge Process for Carbon Oxidation, Nitrification and Denitrification*, BBS Engineering, Source Link: http://www.bbsengineering.com/portfolio/alternative.php3, January (2005).
26. G. T. Daigger, M. H. Jr. Robbins, and B. R. MarshalL, The design of a selector to control low F:M filamentous bulking, *Journal of Water Pollution and Control Federation*, 57, 3 (1985).
27. D. F. Bishop, J. A. Heidman, and J. B. Stamberg, Single-stage nitrification-denitrification, *Journal of Water Pollution Control Federation*, 48, 3, 520–532 (1976).
28. M. C. Goronszy, and J. White, Biological nutrient removal with sludge bulking control in a batch activated system, *Proceedings of the 14th Biennial International Association on Water Pollution Control and Research Conference*, Brighton, UK (1988).
29. J. H. Rensik, New approach to preventing sludge bulking, *Journal of Water Pollution Control Federation*, 46, 1888 (1974).
30. S. R. Linne, and S. C. Chiesa, Operational variables affecting performance of the selector - complete mix activated sludge process, *Journal of Water Pollution and Control Federation*, 59, 7 (1987).
31. J. Chudoba, et al., Control of activated sludge filamentous bulking: I. effect of hydraulic regime or degree of mixing in an aeration tank, *Water Research* 7, 1163 (1973).
32. J. Chudoba, et al, Control of activated sludge filamentous bulking: II. selection of microorganisms by means of a selector, *Water Research*, 7, 1389 (1973).
33. J. Chudoba, et al, Control of activated sludge filamentous bulking: III. effect of sludge loading, *Water Research*, 8, 231 (1974).
34. G. T. Daigger, and G. A. Nicholson, Performance of four full-scale nitrifying wastewater treatment plants incorporating selectors, *Journal of Water Pollution Control Federation*, 62, 5, 676–683 (1990).

35. L. K. Wang, Chemistry of nitrification-denitrification process, *Journal of Environmental Science*, 21, 23–28, December (1978).
36. Y. D. Lee, E. B. Shin, Y. S. Choi, H. S. Yoon, H. S. Lee, I. J. Chung, and J. S. Na, Biological removal of nitrogen and phosphorus from wastewater by a single-sludge reactor, *Environmental Technology*, 18, 10, 975–986, October (1997).
37. L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.), *Biosolids Treatment Processes*, Humana Press, Totowa, NJ (2007).

LIST OF APPENDIXES

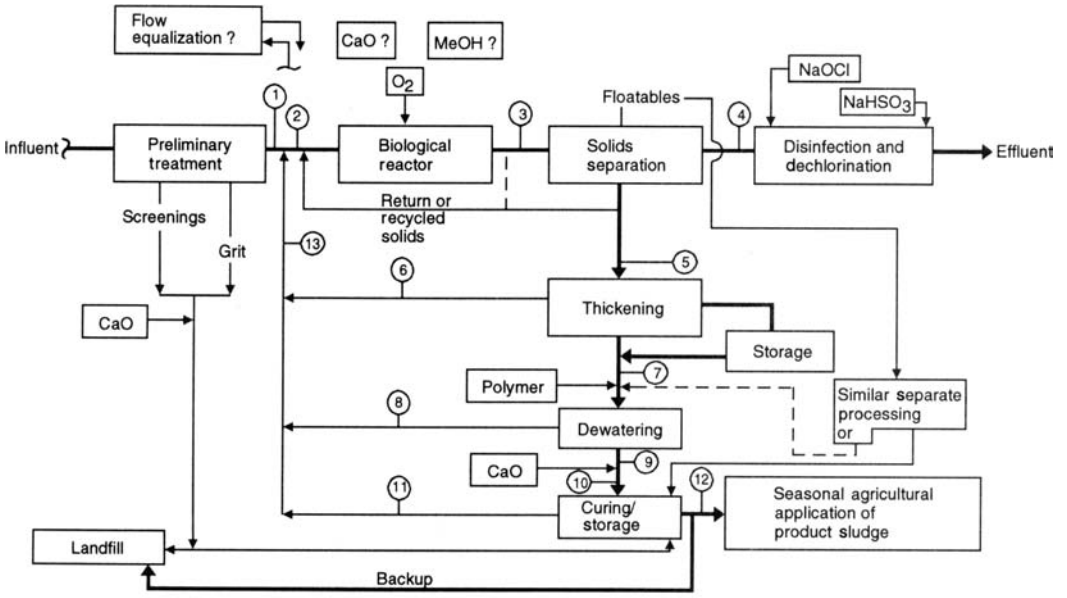
Appendix A

Basic Schematics of Plants A and B Wastewater Treatment Facilities (1)



Appendix B

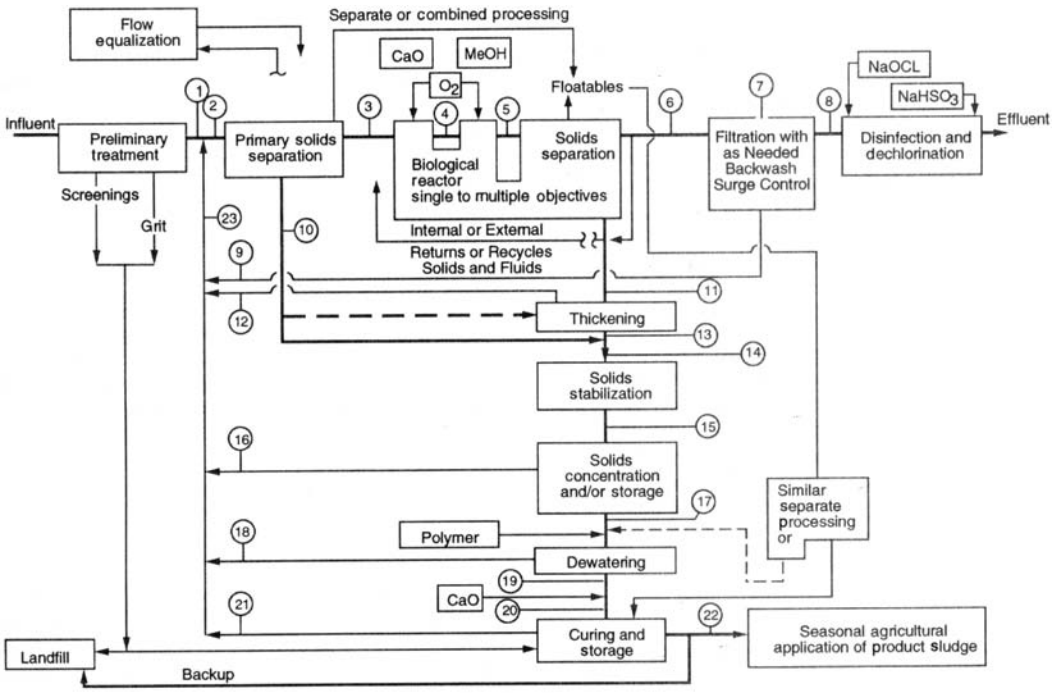
Detailed Schematic of Plant A Wastewater Treatment Facility (1)



Circled numbers indicate process points for mass balance calculations. Note also that schematic is not applicable to attached growth systems, because they lack all forms of primary treatment.

Appendix C

Detailed Schematic of Plant B Wastewater Treatment Facility (1)



Circled numbers indicate process points for mass balance calculations.

Appendix D

Mass Balance for Plant B (1)

	mg/L Equivalents															Normalized Flow[a]	
	SS			$CBOD_5$			COD			Phosphorus as P			Nitrogen as N			Actual mg/L SS	Flow fraction of Q
Processing point	ISS	VSS	TSS	Sol.	Part.	Tot.	Sol.	Part.	Tot.	Sol.	Part.	Tot.	Sol.	Part.	Tot.		
1. Raw influent	52	123	175	50	100	150	100	180	280	5.0	1.0	6.0	24.0	6.0	30.0	175	1.00000
23. Recycles[b]	15	22(26)	37(41)	3	3	6	6	32	38	0.4	0.3	6.7	2.4	1.9	4.3	330	0.11259
2. Total influent	67	145(159)	212(216)	53	103	156	106	212	318	5.4	1.3	6.7	26.4	7.9	34.3	190	1.11259
3. Prim. effluent	25	55(59)	80(84)	53	44	97	106	81	187	6.4	0.6	6.0	26.4	3.1	29.5	72	1.10995
4. Rx effluent[c]	25	58	83	8	20	28	30	81	111	4.9	1.1	6.0	24.1	5.4	29.5	75	1.10995
5. Rx effluent[d]	25	42	67	2	6	8	20	59	79	5.2	0.8	6.0	25.7	3.8	29.5	60	1.10995
6. Sec. effluent	6	9	15	2	1	3	20	13	33	5.2	0.2	5.4	25.7	0.8	26.5	14	1.09955
7. Fi. Rx effi.[e]	6	14	20	2	2	4	20	20	40	5.1	0.3	5.4	20.2	1.3	21.5	18	1.09955
8. Final effluent	1	4	5	2	—	2	20	5	25	5.1	0.1	5.2	20.2	0.3	20.5	5	0.99955
9. Backwash	5	10	15	—	2	2	—	15	15	—	0.2	0.2	—	1.0	1.0	150	0.10000
10. Prim. sludge	42	90	132	—	59	59	—	131	131	—	0.7	0.7	—	4.8	4.8	50,000	0.00264
11. Was. sec. solids	19	33	52	—	5	5	—	46	46	—	0.6	0.6	—	3.0	3.0	5000	0.01040
12. Thk overflow	1	2	3	—	—	—	—	2	2	—	—	—	—	0.2	0.2	380	0.00795
13. Thk sec. sl.	18	31	49	—	5	5	—	44	44	—	0.6	0.6	—	2.8	2.8	20,000	0.00245
14. Sl. to Dig	60	121	181	—	64	64	—	175	175	—	1.3	1.3	—	7.6	7.6	35,600	0.00509
15. Sl. alt. Dig	60	72(77)	132(137)	3	9	12	7	105	112	0.4	0.9	1.3	2.6	5.0	7.6	25,900	0.00509
16. Supernatant	6	7(9)	13(15)	1	1	2	3	10	13	0.2	0.1	0.3	1.1	0.5	1.6	6200	0.00211
17. Sl. to dewater	54	65(68)	119(122)	2	8	10	4	95	99	0.2	0.8	1.0	1.5	4.5	6.0	40,000	0.00295

18. Filtrate	3	3(5)	6(8)	2	—	2	3	5	8	0.2	—	0.2	1.3	0.2	1.5	2400	0.00253
19. Cake	51	62(63)	113(114)	—	8	8	1	90	91	—	0.8	0.8	0.2	4.3	4.5	250,000	0.00045
20. Limed cake	79	62(63)	141(142)														
21. Runoff	Not considered in this example																
22. Cake to land	Same as Processing Point 16 in this example																

[a] To find the actual flow rates, multiply the sludge processing point mg/L equivalents by plant flow and divide by expected actual sludge concentration. Liquid stream recycle flows are determined by difference, with their actual pollutant concentrations determined by dividing the mg/L equivalents by the flow fraction.

[b] Recycles is the sum of processing streams 9, 12, 16, and 18; values in parentheses reflect dissolved solids as a result of anaerobic digestion plus the SS.

[c] Intermediate point of reactor effluent with solids residence time of 2 days, see text for additional detail.

[d] Average day oxygen demand, assuming no natural denitrification, can be calculated from the mass balance as follows:

Carbonaceous Oxygen Demand = Processing Point 3 minus 5 = 187 − 79 = 108 mg/L

Nitrogenous Oxygen Demand = 4.6 mg O_2/mg N oxidized = 4.6 [25.7 − (1.0 + 0.5)] = 4.6(24.2) = $\frac{111\text{ mg/L}}{219\text{ mg/L}}$

Total Oxygen Demand, ignoring effluent DO as too small to be significant = 219 mg/L If this was an activated sludge system, for each million L/d flow:

- The average day air supply (ignoring mixing requirements) and assuming 10% oxygen transfer efficiency and 0.28 kg oxygen/m^3 air, is
= [(219 mg/L)(1.0 ML/d)] + [(0.10 efficiency)(0.28)(1,440 min/d)(60 s/m)] = 0.09 m^3/s (or 191 cfm)
- and the MLSS in an 18-h detention time aerator, for a solids residence time of 15 days is
= (15 d)(67 mg/L/d)(24 h/18 h) = 1340 mg/L

[e] Filter Reactor Effluent reflects net result of methanol addition to achieve 5 mg/L NO_3-N removal, to a process stream containing 3 mg/L DO, where:
- Methanol COD = 1.5 [DO + 2.9(NO_3-N)] = 1.5 [3 + 2.9(5.0)] = 1.5(17.5) = 26 mg/L
- Synthesized Solids - 0.25 (COD added) = 0.25 (26) = 5 mg/L VSS
- Denitrification in this example will yield about (3.6 mg $CaCO_3$/mg NO_3-N)(5 mg/L NO_3-N) = 18 mg/L $CaCO_3$ alkailinity

Appendix E

Mass Balance for Plant B (1)

	mg/L Equivalents[a]															Normalized Flow[b]	
	SS			$CBOD_5$			COD			Phosphorus as P			Nitrogen as N			Actual mg/L SS	Flow fraction
Processing point	ISS	VSS	TSS	Sol.	Part.	Tot.	Sol.	Part.	Tot.	Sol.	Part.	Tot.	Sol.	Part.	Tot.		of Q
1. Raw Influent	52	123	175	50	100	150	100	180	280	5.0	1.0	6.0	24.0	6.0	30.0	175	1.00000
13. Recycles[c]	6	6	12	—	2	2	—	10	10	—	0.1	0.1	—	0.5	0.5	500	0.02404
2. Total Influent	58	129	187	50	102	152	100	190	290	5.0	1.1	6.1	24.0	6.5	30.5	183	1.02404
3. Reactor eff.[d]	58	80	135	2	12	14	20	117	137	4.9	1.2	6.1	24.5	6.0	30.5	135	1.02404
4. Final effluent	6	9	15	2	1	3	20	13	33	4.9	0.1	5.0	24.5	0.7	25.2	15	0.99944
5. Waste solids	52	71	123	—	11	11	—	104	104	—	1.1	1.1	—	5.3	5.3	5000	0.02460
6. Solids	3	3	6	—	1	1	—	5	5	—	0.1	0.1	—	0.3	0.3	300	0.01992
7. Underflow	49	68	117	—	10	10	—	99	99	—	1.0	1.0	—	5.0	5.0	25,000	0.00468
8. Filtrate	3	3	6	—	1	1	—	5	5	—	—	—	—	0.2	0.2	1500	0.00412
9. Cake	46	65	111	—	9	9	—	94	94	—	1.0	1.0	—	4.8	4.8	200,000	0.00056

10. Limed cake[e]	74	65	139
11. Runoff	Not considered in this example		
12. Cake to land	Same as Processing Point 10 in this example		

[a] To find the actual kg mass/d, multiply the mg/L equivalent and the plant flow ML/d; lb/d mass is calculated by multiplying mg/L by 8.34 and the plant flow (mgd).

[b] To find the actual flow rates, multiply the sludge processing point mg/L equivalents by plant flow and divide by expected actual sludge concentration. Liquid stream recycle flows are determined by difference, with their actual pollutant concentrations determined by dividing the mg/L equivalents by the flow fraction.

[c] Recycles is the sum of processing streams 6, 8, and 11.

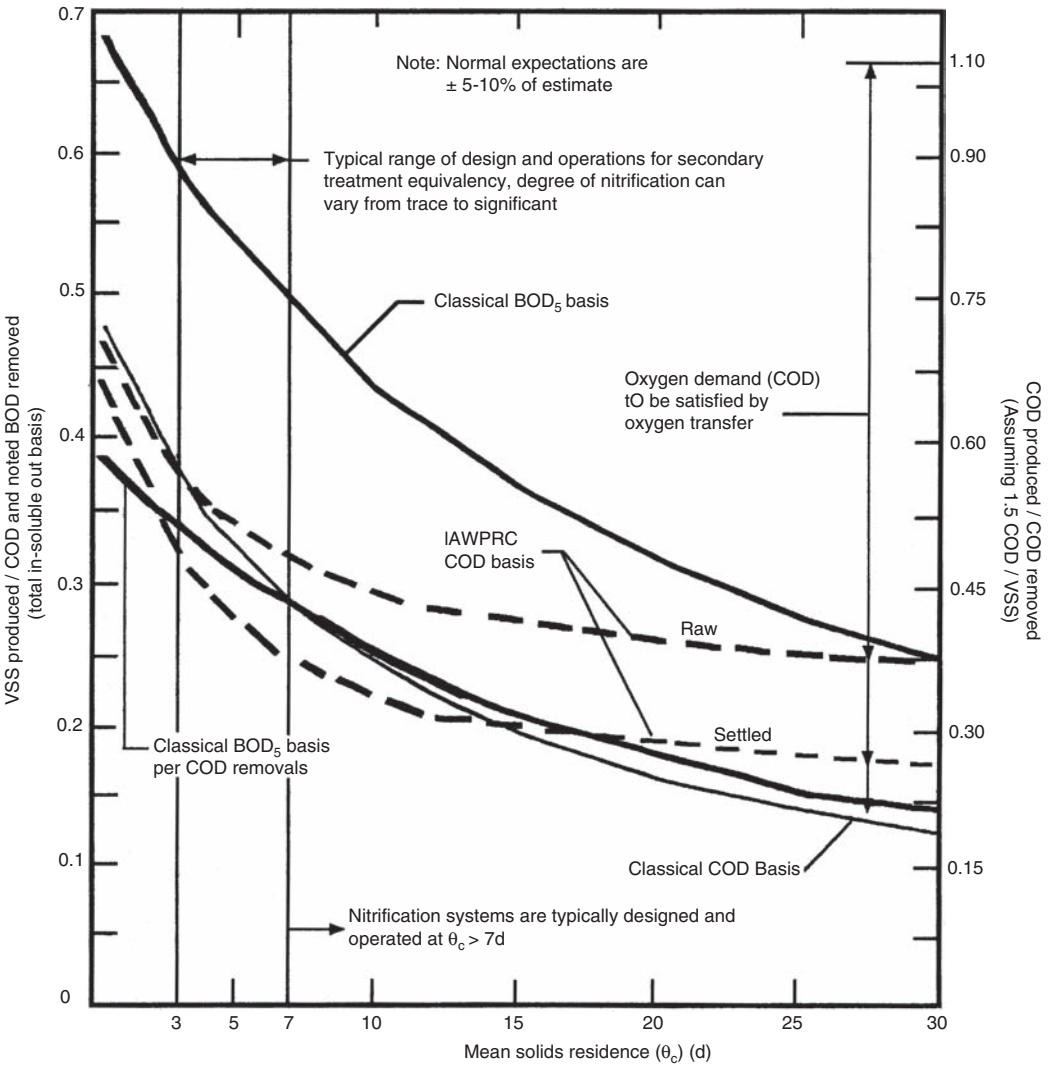
[d] Nonblodegradable VSS were assumed to maintain the characteristics found in the influent wastewater; newly synthesized VSS were assumed to have the following characteristics:

COD/VSS = 1.4; N/VSS = 0.1; P/VSS = 0.02; the BOD:COD ratio was assumed to be 1:10 in the plant effluent and product solids.

- Average day oxygen demand, assuming no natural denitrification, can be calculated from the mass balance as follows: Carbonaceous Oxygen Demand = Processing Point 2 minus 3 = 290 − 137 = 153 mg/L Nitrogenous Oxygen Demand = 4.6 mg O_2/mg N oxidized (Oxidized N is the soluble nitrogen in Processing Point 3 minus soluble nonbiodegradable nitrogen and effluent ammonium); assume nonbiodegradable nitrogen is 1.0 mg/L and effluent ammonium is 0.5 mg/L on the average day = 4.6[24.5 − (1.0 + 0.5)] = 4.6(23.0) = 106 mg/L Total Oxygen Demand, Ignoring effluent DO as too small to be significant = 259 mg/L If this was an activated sludge system, for each million liters/d flow:
  The average day air supply (Ignoring mixing requirements), assuming 10% oxygen transfer efficiency and 0.28 kg oxygen/m^3 air, is = [(259 mg/L)(1.0 ML/d)] + [(0.10 efficiency)(0.28)(1,440 min/d)(60 s/m)] = 0.107 m^3/s (or 226.7 cfm)
  And the MLSS in a 18-h detention time aerator, for a solids residence time of 15 days is = (15 days)(138 mg/L/d)(24 h/18 h) = 2800 mg/L [e] The alkalinity demand for nitrification is:
  (7.1 mg $CaCO_3$/N oxidized) (23.5 mg N oxidized/L) = 167 mg/L, from which, if it is desired to have a residual alkalinity of 50 mg/L as $CaCO_3$, the facility will need a supplemental alkalinity of (167 + 50) − 120 = 97 mg/L as $CaCO_3$, which corresponds to a lime (CaO) dose of 97 + 1.8 = 54 mg/L as CaO
- CaO added at 25% of dry weight mass, which, for this example, was assumed satisfactory for PFRP (Processes for Further Reduction of Pathogens) requirements.

Appendix F

Volatile Solids Production (1)



Appendix G

Design Examples: Influent Wastewater Peaking Factors (1)

Appendix G
Design Examples: Influent Wastewater Peaking Factors (1)

Condition	Percent of time conditions[a]	Ratio of noted condition to average day pollutant mass			
		Flow	SS and organics	Total P and N	Matching alkalinity
Minimum month	7.7	0.7	0.8	0.8	0.9
Average day	50.0	1.0	1.0	1.0	1.0
Maximum month	92.3	1.5	1.3	1.2	1.1
Maximum week	98.1	1.9	1.6	1.4	1.3
Maximum day	99.7	2.5	2.1	1.7	1.5
Maximum hour	99.99	3.0			

[a] Equivalent percent of time conditions are less than or equal to stated values.

9
Column Bioreactor Clarifier Process (CBCP)

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CONTENTS

Abstract A novel columnar bioreactor system comprising an integrated bioreactor clarification system with layers of pseudoliquified activated sludge was demonstrated at a pilot scale for the treatment of municipal wastewaters. The system was mechanically simple, avoiding the need for any mechanical components for biomass recirculation, mixed liquor recirculation, and sludge collection and wastage. Wastewater flows within the different system zones were achieved by gravity only. The system achieved very high carbon, nitrogen, and suspended solids removal efficiencies with the final effluent quality approaching that of tertiary effluents with respect to BOD, TSS, ammonia, nitrates, and total nitrogen at very low HRTs of 4.8 hours. Despite the absence of enhanced biological phosphorous removal, with chemical addition the system was able to consistently achieve final effluent TP concentrations of about 0.7 mg/L.

Key Words Activated sludge • BOD • column bioreactor • nitrogen • phosphorus • tertiary effluent • TSS.

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_9

1. BACKGROUND

The effects of pollution have been recognized as an urgent problem in many areas of the world. Systems used for wastewater treatment and purification are not necessarily highly efficient. The problem of eutrophication in surface water sources directly connected with nitrogen and phosphorus pollutants is still a predominant concern in most of the countries of the world (1–5, 7).

The common design of conventional primary and secondary treatment systems uses a water depth up to 5 m. This results in shallow aeration tanks, low concentration of activated sludge, separate clarifiers with low hydraulic loading, recirculation pumping facilities, and interconnecting piping, which require large areas of land and high capital costs. In particular locations, highly productive agricultural lands or expensive urban lands are used to accommodate these facilities.

As an alternative, the Ukrainian National Research and Development Institute for Municipal Facilities and Services in Kiev has investigated and developed technologies for complete biological treatment of wastewaters in deep reactor vessels (8–10). Recognizing the global and regional needs for improving wastewater treatment, specifically with respect to nitrogen and phosphorus removal from inland waters, the Institute has been working to further develop them with input from Canada's Programs of Industrial Collaboration, Professional Partnerships and Environmental Management in Ukraine, Conestoga-Rovers and Associates and CRA, Waterloo, Ontario, Canada.

This chapter is based on data derived from several operating facilities and a detailed pilot study with the objective of improving understanding of the fundamental processes in this technology and its limitations for potential use in the world.

2. INTRODUCTION

Continuous monitoring and professional experience indicate that the quality of natural inland waters, including the Dniper River, is deteriorating. This is a result of untreated or insufficiently treated municipal, industrial, and agricultural wastewater being discharged into the rivers. The deterioration is also attributed to storm and snowmelt runoff from agricultural and urban lands.

The aquatic environment is being destabilized and deteriorated by the ever increasing quantities of discharged mineral and biological pollutants. The industrial and municipal wastewater, which is treated, does not meet environmental standards. Waste parameters that exceed the "acceptable limit discharge" (ADL) include: oily wastes, nitrogen, phosphorous, suspended solids, phenol, pesticides, and heavy metals. Most industrial wastes are either untreated or are discharged from treatment plants that do not adequately perform or are grossly overloaded. These types of industry include: mineral or organic fertilizer; grain, meat, or dairy processing; metallurgical and cooking plants; chemical and oil production.

Much of this industrial discharge flows through municipal treatment plants, which are not designed for the removal of such contaminants, and directly into the receiving waters of the Dniper and its tributaries. Some municipal discharge by-laws set "acceptable limit concentrations" (ALC), but these are continuously and significantly exceeded, as are the effluent

standards (30 to 40 mg/L, BOD, and TSS) set for the municipal discharge to the receiving waters. Data indicates that most treatment facilities are overloaded or receive specialized contaminants for which they are not designed. This situation results in inadequate treatment of the discharge to the Dniper River basin. Stringent effluent quality criteria are currently under consideration for discharges to the Dniper River. Such criteria would stipulate ultimate BOD, ammonia-nitrogen (NH_3-N), and total phosphorous (TP) concentrations of 3.0, 0.39, and 1.0 mg/L, respectively.

Although municipal and industrial wastewater treatment technologies are well developed, capital and operational costs of treatment processes are high and land requirements in some areas may be extensive.

Experience indicates that multistage biological or combined biological and chemical treatment technologies with high-density activated microorganisms and fixed-film media improve treatment efficiency of most control parameters. Developing experience with the application of cost-effective novel treatment technologies that achieve nitrification and denitrification and tertiary treatment by biosorption is the key to abatement of river pollution.

3. DESCRIPTION OF NOVEL TREATMENT TECHNOLOGY

3.1. Concepts of Biological Processes

The unique concepts inherent in this new technology are based on increasing concentration of active biomass and decreasing biomass specific loading and oxidation rate, as well as accommodation of microbial treatment and solid-liquid separation in a single reactor. These concepts create and maintain more favorable conditions for the bacterial cultures during the active respiration time in the treatment system.

Encompassed within a single volume are the removal processes of biosorption and organic oxidation as well as simultaneous separation of treated water by filtration through the pseudoliquified layers of activated sludge and fixed-film microorganisms. The key features of this technology are the incorporation of biological oxidation of organics, nutrients removal (nitrogen and phosphorous), and clarification in a single deep reactor vessel. Consequently, the concept of a single reactor, in which these processes are employed, minimizes the total volume within which complete mixing and retention of biomass occurs.

Low energy vortexes are used to create alternating layers of low and very high concentrations of activated sludge that simultaneously increase active biomass concentration and ensuing biodegradation as well as provide a physical filtration medium. Vortexes of pseudoliquified layers are created by using the energy of injected air or pumping sludge–water mixtures and recirculated flows. These methods ensure the stable existence of pseudoliquified layers of activated sludge despite changes in the quantity of wastewater and even during the complete absence of flowing water within the structure.

The increased exchange of recirculation flows between aeration and recirculation zones far exceeds those occurring between conventional aeration tank-clarifiers. Because the aeration zone acts as a dissolved oxygen distribution system, especially at high sludge recirculation flows, the volume of the aeration zone is lower than that of the clarification zone, where the majority of the biological processes occur. This system is more efficient than the conventional

activated sludge wastewater treatment, employing common aeration tank and clarifier configuration. The bioreactor operates as a combination of suspended growth and fixed-film system and is protected by patents (Nos. 7903522, 7907418, 604287, 1530573, 1638122, 1721028, 2451892) (10–16).

Schematic of compact reactor-clarifier of a single column-type with central aeration zone is shown in Fig. 9.1. The columnar reactor is designed as vertical column, divided by the concentrically cylindrical and conic partition systems into aeration zone, degasification and recirculation zones, and clarification zone with pseudoliquified layers of activated sludge. The aeration, degasification, recirculation, and clarification zones are hydraulically connected with each other by the recirculation channel system. The principal distinctive features of the column-type reactor are: creation of hydrodynamic conditions, which are determinate of intensive recirculation flows and mass exchange between aeration and clarification zones and creation of pseudoliquified layers of activated sludge on the column height owing to the energy of recirculation flows. Thus, creation of pseudoliquified layers of activated sludge in the column-type reactor is not dependent on the influent flow (i.e., without influent flow) as in the conventional aeration tank-clarifier system.

In the column-type aeration tank, the pseudoliquified layers of activated sludge are used as a reactor for wastewater pollutants oxidation. Intensification of wastewater pollutants oxidation processes is achieved by the high concentrations of activated sludge (4,000 to 8,000 mg/L), enhanced aerobic conditions in whole-reactor volume as a result of high recirculation rates, and retention of biomass and particulates as a result of filtration through the pseudoliquified layers of activated sludge.

Mass transfer, effectiveness of aeration, biosorption, and oxidation processes are increased because of creation of recirculating aeration system, counter-flows, high pressure, recirculation between aeration and clarification zones, and use of pseudoliquified layers as an oxidation reactor.

Column aeration tank-clarifier design pertains to the combined aeration facilities with the combination of aeration processes, recirculation, oxidation of pollutants into the pseudoliquified layers of activated sludge, and simultaneous clarification of activated sludge and treated wastewater owing to filtration of wastewater through pseudoliquified layers of activated sludge. Filtration rate of wastewater through pseudoliquified layers of activated sludge is low due the high filtration surface as a result of increasing number of layers.

3.2. Distinction of Biosorption and Oxidation Processes in the Pseudoliquified Activated Sludge Bioreactor

The main factors affecting activated sludge processes intensification are loading and hydraulic parameters. As the activated sludge concentration increases, conditions that are conducive for microbial activity are created, including increased rates in specific biosorption and biochemical oxidation, enhanced mass transfer, clarification, and oxygen transfer.

Reactors with pseudoliquified layers of reacting substances are widely used in the chemical industry, as well as in natural water and wastewater ecosystems. Experience indicates that the rate of oxidation processes in the pseudoliquified layer is several times higher than in a completely mixed reactor (17, 18).

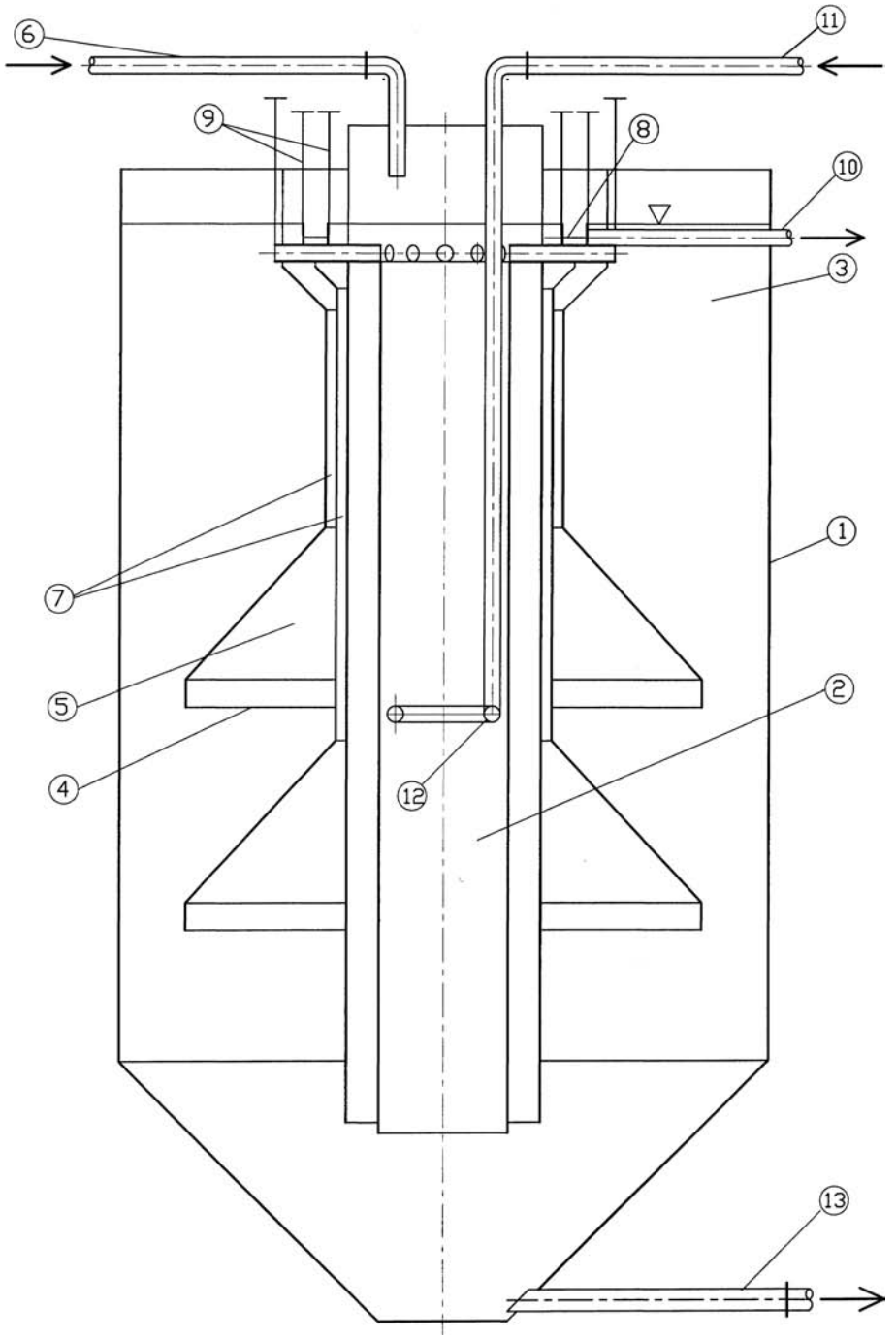

Fig. 9.1. Schematic of aeration tanks-clarifiers of single column-type with central aeration chamber.

Vortex pseudoliquified layers of activated sludge, which are formed in the column aeration tank-clarifiers, are not influenced by influent wastewater flows, and are distinctively more stable than the other kinds of pseudoliquified layers, such as: upflow, direct flow and counter-flows, and spout flows.

Activated sludge in the pseudoliquified layers consists of biological flocs that are matrices of micro-organisms. Biological flocs of activated sludge are maintained in a suspended state by the hydrodynamic pressure of vortex flows and up-flow of liquid, forming a heterogeneous system with the water (pseudoliquified layer).

Biological flocs of activated sludge are continuously moving (circulated) and create local thickening and stabilization zones in an "upper layer." The pseudoliquified layer is in a dynamic equilibrium state governed by the upflow velocity and size of biological flocs. The creation of pseudoliquified layers is influenced by a differential density of activated sludge and wastewater, as well as the size of biological flocs.

Biological flocs of activated sludge come into contact with each other and agglomerate to form a "suspended filter." Aerated wastewater is filtered through that suspended filter of activated sludge and come into contact with biological flocs while colloidal particles and suspended solids are removed owing to biosorption processes (19, 20).

In addition to mixing and layered homogeneity, the presence of dissolved oxygen in the pseudoliquified layers of activated sludge is necessary for biosorption and oxidation processes. Dissolved oxygen presence in the pseudoliquified layers of activated sludge is achieved by the intensive recirculation rate between aeration column and clarification chamber that is up to 800% to 1.000% of the influent wastewater flow.

Thus, the pseudoliquified layers of activated sludge serve as a reactor for simultaneous biosorption and oxidation processes in view of the following distinctive features:

1. Large flocs surface area, intensive contact processes, turbulence, and filtration processes.
2. Simultaneous biosorption and clarification of activated sludge and treated wastewater.
3. Enhanced oxidation processes as a result of the long solids residence time (SRT) in the system.
4. Creation of stabilized vortex pseudoliquified layers of activated sludge by the energy of directional aeration/recirculation flows, at each conical structure.
5. Improved oxygen and substrate transfer between the aeration column and clarification chamber.

3.3. Process Configuration

The Institute has employed various types of engineering designs and configurations in developing these technologies. These designs incorporate aeration tanks divided into zones of aeration and clarification by vertical and inclined partitions. The zones are hydraulically connected and include zones of degasification, recirculation, and zones of suspended layers of activated sludge (21, 22).

Two main designs of this technology are column-type and gallery-type configuration. Aeration tanks of columnar type are either developed within the main reactor or as a separate central column surrounded by multiple-column reactor-clarifiers. Column-type pseudoliquified activated sludge bioreactors (PASB) use circular column-type or polyhedral reactor and

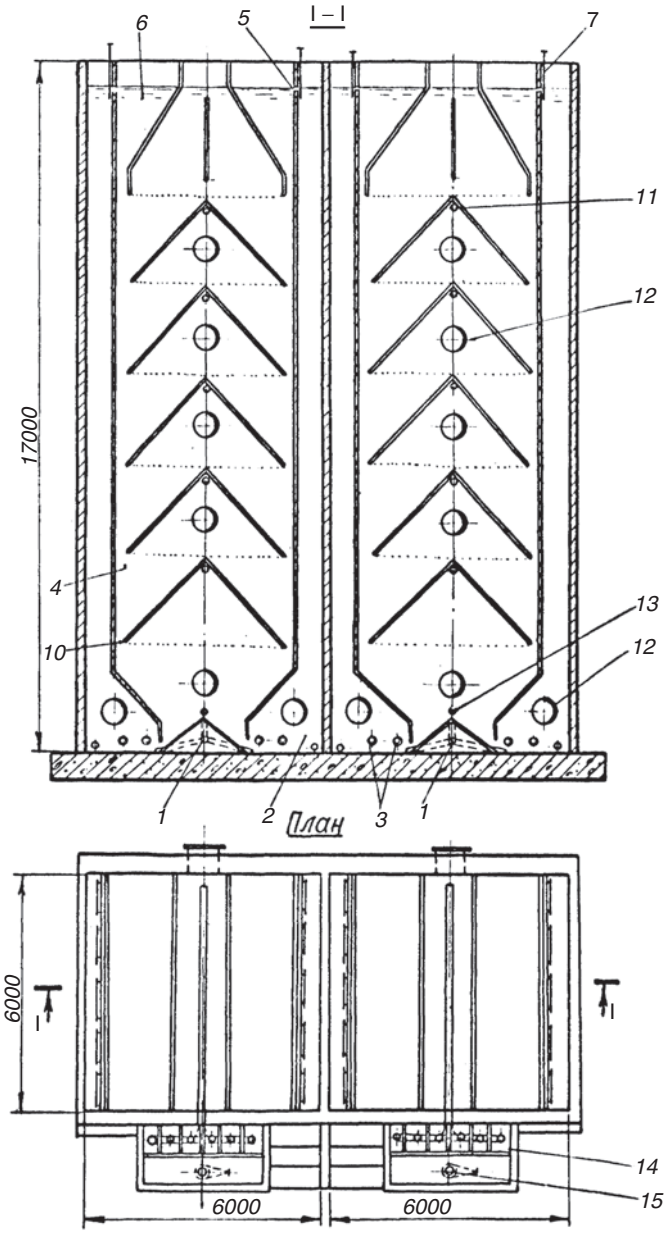

Fig. 9.2. Layout of a single column-type with the external aeration chamber.

range from 1.8 to 12 m in diameter and with a height of up to 20 m. Figure 9.1 shows a schematic of aeration tanks-clarifiers of single column-type with a central aeration chamber, whereas Fig. 9.2 shows a layout of single column-type with an external aeration chamber.

The columnar aeration reactor consists of the following parts (see Fig. 9.1): vertical column reactor frame; aeration chamber; degasification chamber; recirculation chamber; pseudoliquified layers of activated sludge zone; treated wastewater protected chamber; inlet pipe; treated wastewater collection pipes; circular treated wastewater collection chamber; moving windows with a 90° weir treated wastewater collection chamber for treated water equalization; outlet pipe; air supply pipe; diffuser; concentric partitions that separate the aeration and clarification zones; recirculation pipe in the aeration zone; recirculation pipes of circulation system between an aeration and clarification zones; flow gates on the recirculation pipes; cone partitions of the pseudoliquified layers of activated sludge; recirculation channel between aeration and clarification zones; fixed media; jet aeration extraction and discharge pipe. Two recirculation systems exist in the column aeration reactors: within the aeration chamber and between aeration and clarification chambers.

After primary treatment, wastewater enters the aeration column, where it contacts high levels of dissolved oxygen mixed with activated sludge owing to jet mixing of air and wastewater from the bottom of the reactor vessel. Oxygen-rich sludge-wastewater is discharged by the recirculation pipes into clarification chamber as a result of density liquid levels differential between the chambers. The top of the main reactor vessel acts as a degasification chamber from which treated wastewater is drawn. Flow then occurs by gravity, with most wastewater recirculating to enter the bottom of the aeration system through the recirculation channel. The energy of the aeration/recirculation flows is used to create the low energy vortexes, and a dense layer of activated sludge forms at each conical structure.

The treated wastewater is filtered through the pseudoliquified layers of activated sludge, and flows into a treated water protection chamber, where it is discharged via the treated wastewater chamber. Equalization of treated water is achieved by moving windows with a 90° weir in the circular treated wastewater collection chamber for hydraulic loss compensation in the collection pipes from the layers.

Moreover, the columnar aeration tank may be equipped with media for fixed-film nitrification and denitrification and increase of active biomass concentration in the reactor. Consequently, suspended activated sludge microorganisms and fixed-film microorganisms exist in the reactor. Both take part in carrying out simultaneous processes of biological nitrification and denitrification and filtered clarification of activated sludge and treated water. Treated wastewater is circulated several times by the aeration and recirculation system through fixed media and pseudoliquified layers of activated sludge, thus, increasing the contact time of pollutants with microorganisms and improving water quality.

Due to the depth and the resulting high biosolids concentration and increased oxygen transfer efficiencies, columnar bioreactors can reduce land requirements by up to 80%. Also, a high average of MLSS concentration translates to faster kinetics of BOD and nitrogen removal.

Column-type systems are presently used by small- to medium-sized municipalities, mining camps, health resorts, and industrial facilities such as dairy, meat, poultry, and fish processing plants. The modular treatment systems can be constructed in sizes up to $50,000\,m^3/day$ (11 MGD). View of a constructed combined column-type wastewater treatment plant is presented on Fig. 9.3. The plant is manufactured from the steel, and includes the primary

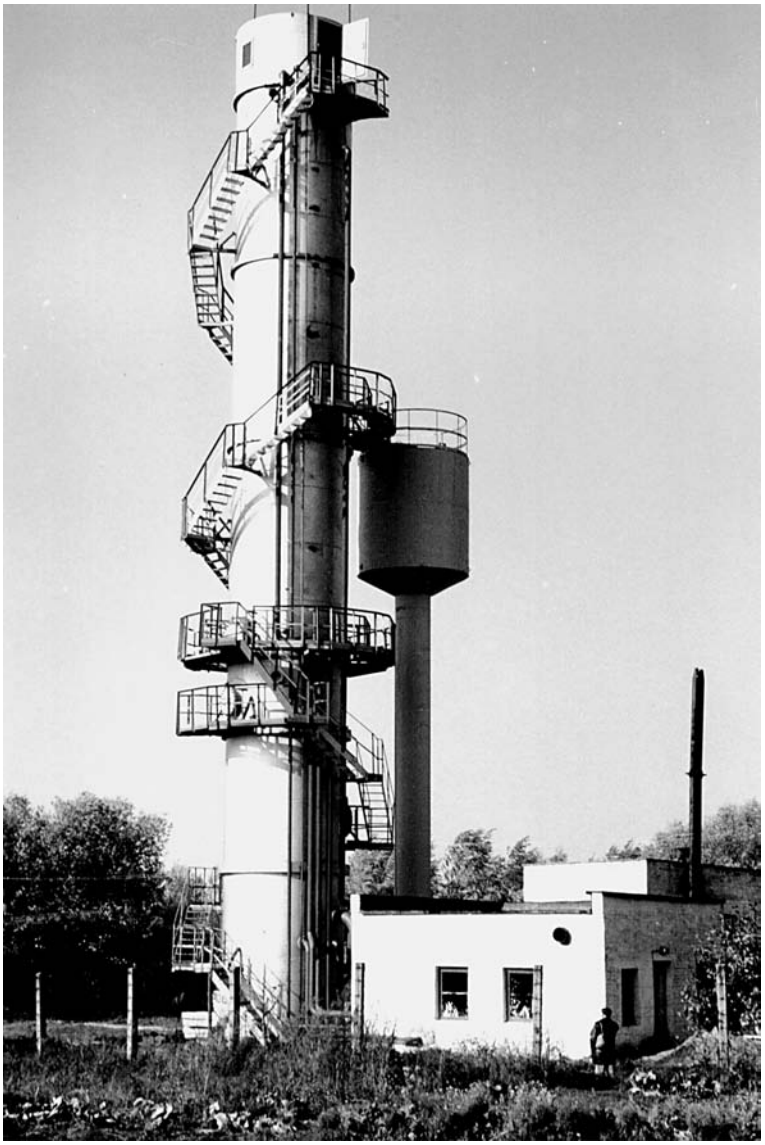

Fig. 9.3. View of combined column-type of wastewater treatment plant manufactured from steel, including the primary settler, column aeration tank-clarifier, tertiary filter with the gravel media, capacity—700 m^3/day.

settler, column aeration tank-clarifier, and tertiary filter with the gravel media, with a capacity of 700 m^3/day. A similar plant with a capacity of 22, 000 m^3/day, illustrating the modular construction of this technology is presented in Fig. 9.4.

Fig. 9.4. View of actual combined column-type design of wastewater treatment plant manufactured from steel, including the primary settler, column aeration tank-clarifier, tertiary filter with the gravel media, capacity—22, 000 m^3/day.

On the other hand, Fig. 9.5 shows a typical design of gallery-type system employing rectangular tanks with a depth of 5 to 6 m. The reactor-clarifiers of the gallery-type feature one or two clarification zones, followed by a secondary clarification chamber with one or two high concentration layers of activated sludge. Ranging in sizes from 2,000 to 400, 000 m^3/day (110 MGD), the gallery-type systems are used by small, medium, and large municipalities, as well as by various industries. These systems require 20% less area compared to conventional activated sludge plants.

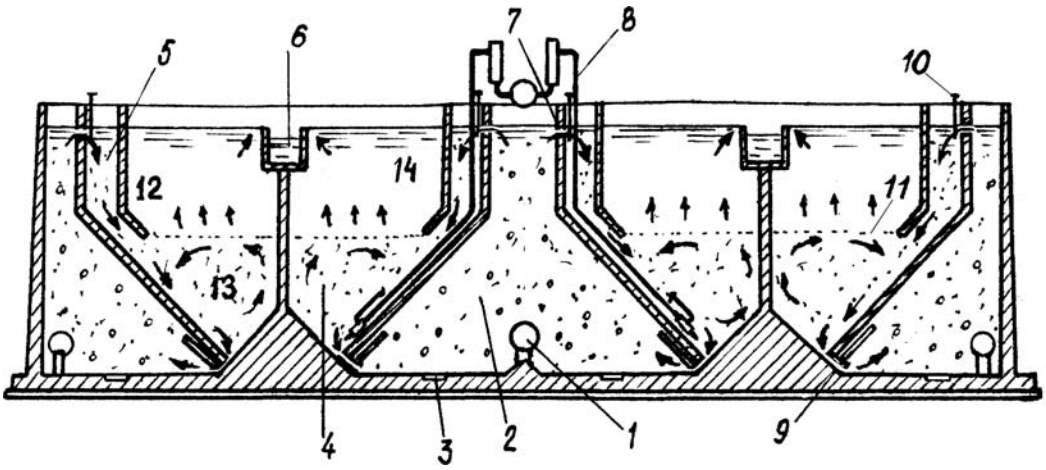

Fig. 9.5. The typical design of gallery-type of aeration tank-clarifier system employing rectangular tanks.

The plants as described above use low-pressure diffused or jet aeration systems. Higher oxygen transfer efficiency and substrate utilization rate in the aeration zones of bioreactor is achieved by specific air distribution under countercurrent and concurrent conditions. The air or air-wastewater mixture is introduced at no more than 5 m below the water surface. Because of the special recirculation partitions, the air bubbles are spread over the entire depth of the aeration tanks. This arrangement improves oxygen transfer efficiency as well as substrate utilization rate and recirculation in both the aeration zones and between the aeration and clarification zones. The units can be manufactured from steel, stainless steel, ferrocement, fiberglass-cement, and fiberglass-plastics in modules ready for shipment and installation either assembled or assembled on site. This method ensures the maximum possible compactness and modular construction.

3.4. Operating Process Parameters

The results of research and operational experience have demonstrated that the rates of biosorption and oxidation processes are up to twice as efficient as those of conventional systems. This appears to be owing to favorable conditions for microorganisms, the increase in mass exchange and the prolonged contact within the pseudoliquified layers during the filtration of treated wastewater through the flocculent mass of activated sludge.

The results of pilot and full-scale investigations of the above-described biological treatment, including tertiary treatment of municipal and industrial wastewaters, are summarized in Table 9.1. Testing parameters data were calculated as follows:

Table 9.1
Results from various types of the aeration tank-clarifier plants

	Units	Aeration tank-clarifier effluent for stages with HRT, hours, and SRT—sludge age, day							
	HRT, hours	2.5		3		4		6	
Parameters	SRT—sludge age, day	4.4	13	4.7	9.1	13.8	22	32	43
1. COD	mgO_2/L	80.6	58.4	65.0	34.0	58.2	52.7	46.5	49
2. BOD_5	mgO_2/L	7.2	5.5	6.5	3.96	3.5	3.8	6.2	4
3. TSS	mg/L	12.1	11.2	10.5	10.1	9.5	9.7	9.0	8.1
4. $N - NH_4^+$	mgN/L	2.42	2.58	0.61	0.43	0.4	0.78	0.2	0.08
5. $N - NO_2^-$	mgN/L	0.41	0.335	0.27	0.209	0.21	0.19	0.006	0.017
6. $N - NO_3^-$	mgN/L	0	0	0	0	0	0.35	0	0
7. pH	–	7.4	7.5	7.4	7.35	7.4	7.5	7.3	7.3
8. ORP	mV	157	120	135	123	125	110	92	80
9. DO	mgO_2/L	1.8	1.7	1.9	1.7	2.0	1.9	2.5	2.3
10. TMLSS	mg/L	3450	7500	2800	5900	5800	6100	5900	7060
11. TMVSS	mg/L	2415	5250	1960	4130	4060	4270	4130	4942
12. ρ_{org}	$mgBOD_u$/(g·h)	21.2	8.2	22.7	13.4	8.4	8.7	5.4	4.3
13. ρ_{org}	mgCOD/(g·h)	38.5	15.6	37.4	25.1	16.0	16.6	10.0	8.1
14. ρ_{nitr}	mgN_{NH4}/(g·h)	8.0	3.0	5.0	2.5	1.71	2.21	1.3	1.2
15. ρ_{den}	$mgN_{(NO2+NO3)}$/(g·h)	7.88	2.9	4.87	2.47	1.69	2.2	1.23	1.17
16. q_i	mgCOD/(g·day)	924	372	897	602.4	384	398.4	240	196.4
17. q_{nitr}	mgN_{NH4}/(g·day)	192	72	120	60	41	53	31.2	28.8
18. q_{den}	$mgN_{(NO2+NO3)}$/(g·day)	189	69.6	116.8	59.8	40.5	52.8	29.5	28.1
19. OC_i	gCOD/(m^3·day)	2.39	2.1	1.9	2.7	1.67	1.8	1.06	1.04
20. OM_{nitr}	gN_{NH4}/(m^3·day)	0.5	0.4	0.25	0.26	1.78	2.4	1.37	1.52
21. OM_{den}	$gN_{(NO2+NO3)}$/(m^3·day)	0.49	0.39	0.24	0.26	1.76	2.4	1.3	1.48
22. Eff_{nitr}	%	93	90	98	99	97	97	99	99
23. Eff_{den}, %	%	98	98	98	98	99	99	99	98
24. $YieldP_{iyd}$	$g/gBOD_u$	0.31	0.32	0.39	0.38	0.37	0.22	0.26	0.21

The organic nitrogen removal efficiency is determined from the following equation:

$$E_{org} = \frac{N_{org}^{in} - N_{org}^{ex}}{N_{org}^{in}} \cdot 100\% \quad (1)$$

in which: $N_{org}{}^{in}$, $N_{org}{}^{ex}$—organic nitrogen concentrations influent and effluent, respectively, mg/L.

The effectiveness of biological nitrification process can be calculated using the equation:

$$E_{\text{nitr}} = \frac{\left(N_{\text{org}}^{\text{in}} + N_{\text{NH}_4}^{\text{in}}\right) - \left(N_{\text{org}}^{\text{ex}} + N_{\text{NH}_4}^{\text{ex}}\right) - P_i \cdot 0,09}{\left(N_{\text{org}}^{\text{in}} + N_{\text{NH}_4}^{\text{in}}\right) - N_{\text{org}}^{\text{ex}} - P_i \cdot 0,09} \cdot 100\% \tag{2}$$

where $N_{\text{NH4}}{}^{\text{in}}$, $N_{\text{NH4}}{}^{\text{ex}}$ — ammonia-nitrogen concentrations influent and effluent, respectively, mg/L, P_i— yield of activated sludge, mg/L, $P_i \cdot 0.09$—coefficient, that is calculated the part of nitrogen ammonia for bacteria cells growing of activated sludge. The atoms of nitrogen, in accordance with the activated sludge equation —$C_{106}H_{180}O_{45}N_{16}P$, are 9% from the common mass balance.

The effectiveness of the biological denitrification process is obtained as follows:

$$E_{\text{den}} = \frac{N_{\text{tot}}^{\text{in}} - N_{\text{tot}}^{\text{ex}} - P_i \cdot 0,09}{N_{\text{tot}}^{\text{in}} - \left(N_{\text{org}}^{\text{ex}} + N_{\text{NH}_4}^{\text{ex}}\right) - P_i \cdot 0,09} \cdot 100\% \tag{3}$$

where $N_{\text{tot}}{}^{\text{in}}$, $N_{\text{tot}}{}^{\text{ex}}$ — total nitrogen concentrations influent and effluent, respectively, mg/L.

Total nitrogen removal efficiency, E_{tot}

$$E_{\text{tot}} = \frac{N_{\text{tot}}^{\text{in}} - N_{\text{tot}}^{\text{ex}}}{N_{\text{tot}}^{\text{in}}} \cdot 100\% \tag{4}$$

Specific rates of oxidation are calculated on the following equations:

$$\rho_{\text{org}} = \frac{N_{\text{org}}^{\text{in}} - N_{\text{org}}^{\text{ex}}}{a_i \cdot (1 - s) \cdot t} \tag{5}$$

$$\rho_{\text{nitr}} = \frac{\left(N_{\text{org}}^{\text{in}} + N_{\text{NH}_4}^{\text{in}}\right) - \left(N_{\text{org}}^{\text{ex}} + N_{\text{NH}_4}^{\text{ex}}\right) - P_i \cdot 0,09}{a_i \cdot (1 - s) \cdot t} \tag{6}$$

$$\rho_{\text{den}} = \frac{N_{\text{tot}}^{\text{in}} - N_{\text{tot}}^{\text{ex}} - P_i \cdot 0,09}{a_i \cdot (1 - s) \cdot t} \tag{7}$$

$$\rho_{\text{tot}} = \frac{N_{\text{tot}}^{\text{in}} - N_{\text{tot}}^{\text{ex}}}{a_i \cdot (1 - s) \cdot t} \tag{8}$$

Municipal wastewater treatment was carried out at hydraulic retention times (HRT) of 2.5, 3.0, 4.0, and 6.0 hours and biological solids residence time (SRT) of 4.4 to 43 days in gallery-type aeration tank-clarifiers.

The layout and view of a pilot plant of column-type aeration tank-clarifier integrated with tertiary filtration in a single reactor is presented on Figs. 9.6 and 9.7, respectively. The plant includes the primary settler, aeration tank-clarifier, and tertiary filter with the gravel media, which are connected by the pipes. The plant consists of the following facilities: inlet pipe, grit removal tank, pump for pumping of wastewater, primary settler, aeration tank-clarifier, treated wastewater channel, tertiary filter, treated wastewater storage tank for tertiary filter

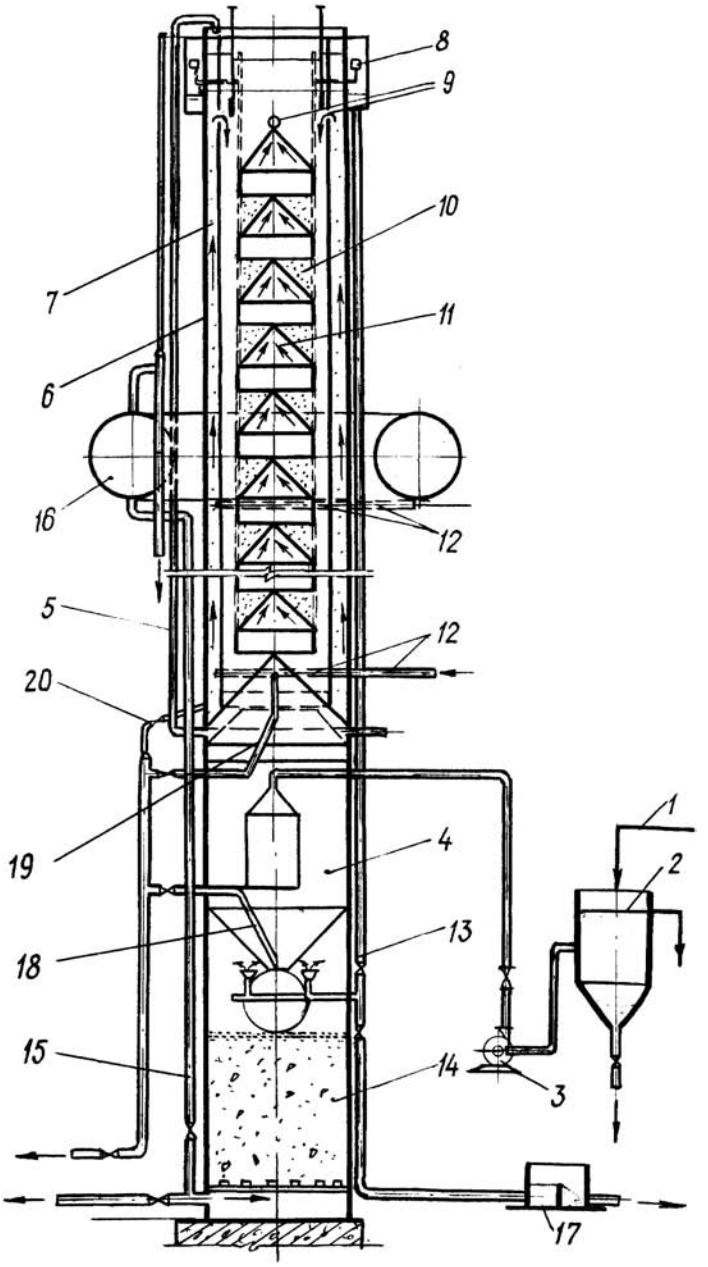

Fig. 9.6. Details of the pilot plant combining columnar aeration tank-clarifier with tertiary filtration in a single reactor.

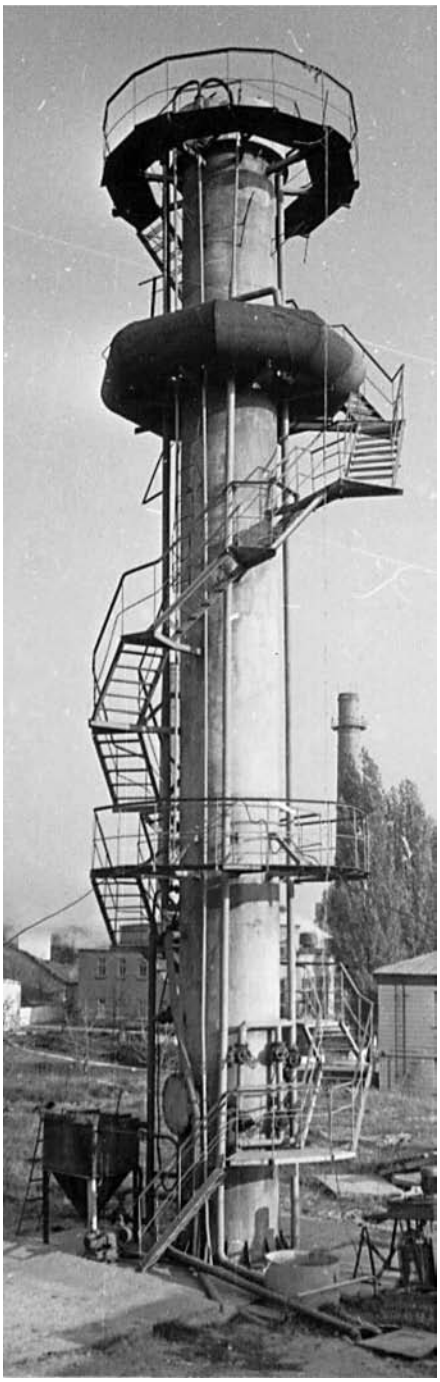

Fig. 9.7. View of the pilot plant combining columnar aeration tank-clarifier with tertiary filtration in a single reactor.

backwashing, outlet pipe, measuring chamber with a 90° weir, waste activated sludge and sludge from the primary settler discharge.

The combined column-type plant is made from a steel pipe with a diameter of 1.6 m. The column aeration tank-clarifier consists of an outside aeration chamber, aeration system with a diffuser depth of 4.3 m, central clarifier chamber with pseudoliquified layers of activated sludge. The volume of primary settler is 5.1 m^3, volume of activated sludge in the aeration tank-clarifier—20.0 m^3, common surface square of pseudoliquified layers of activated sludge—9.35 m^2. The number of tertiary filter sections—2 surface square of each section—1.0 m^2. The volume of tertiary filter backwashing tank is 7.9 m^3.

The integrated bioreactor/clarification/filtration plant works as follows. Wastewater comes into the grit removal tank and is pumped into the primary settler. Settled wastewater from the primary settler is pumped into the aeration chamber of the aeration tank where it contacts activated sludge. In the aeration chamber, wastewater is mixed with an activated sludge and dissolved oxygen. Oxygen-rich sludge water flows through the recirculation pipes into clarification chamber, where oxidation of organic pollutants, clarification of activated sludge, and treated water separation processes occur. Biologically treated wastewater is collected by the chamber and discharged by an outlet pipe into the tertiary filter. The treated wastewater enters the tertiary filter backwashing tank and is discharged through the measuring chamber as final effluent. A major component of the recirculation flow is returned from the clarification chamber into the aeration chamber by the recirculation channel. Primary sludge and waste activated sludge from the aeration tank are discharged by pipe into a measuring tank. The hydraulic loading in the combined column-type plant is 4.63 m^3/h (111.5 m^3/day). Pilot testing data are graphically shown in the Fig. 9.8. The test data of the aeration tank-clarifier of gallery-type, block modular column-type plant, and combined column-type plant are used for drawing correlations and analyses of system kinetics graphs as shown in Figs. 9.8–9.14.

A summary of data obtained over a wide range of operational conditions is presented in Table 9.1. At all times, treatment efficiency was high (see Table 9.1). Average BOD_5, COD, and TSS levels in the final effluent were in the range 3.5 to 7.2 mg/L, 34.0 to 86.0 mg/L, and 8.1 to 12.1 mg/L, respectively. But, at an HRT of 2.5 hours and SRT of 4.4 days, effluent quality parameters were: COD—80.6 mg/L; BOD—7.2 mg/L; TSS—12.1 mg/L; TMLSS—3,450 mg/L; ORP—157.0 mV. After increasing sludge age (SRT) to 13 days, treated water quality improved: COD—58.4 mg/L; BOD—5.2 mg/L; TSS—11.2 mg/L; TMLSS—7,500 mg/L; ORP—120 mV. Nitrification and denitrification processes occurred as reflected by: NH_4-N—2.58 mg/L; NO_2-N—0.335 mg/L; NO_3-N—0.0 mg/L, corresponding to 90% nitrification efficiency and 98% denitrification efficiency.

The HRT was increased further to 3.0 h, SRT—4.7 and 9.1 days, for improving effluent quality criteria. During this period, ammonia-nitrogen oxidation rates were increased: NH_4-N—0.43 mg/L, 0.61 mg/L; NO_2-N—0.209, 0.27 mg/L; NO_3-N—0.0 mg/L, effectiveness of nitrification—98%; denitrification—98% TMLSS—7,350, 7,400 mg/L; ORP—123, 125 mV. Effectiveness of ammonia-nitrogen conversion met required standards, but the nitrites reduction was lower.

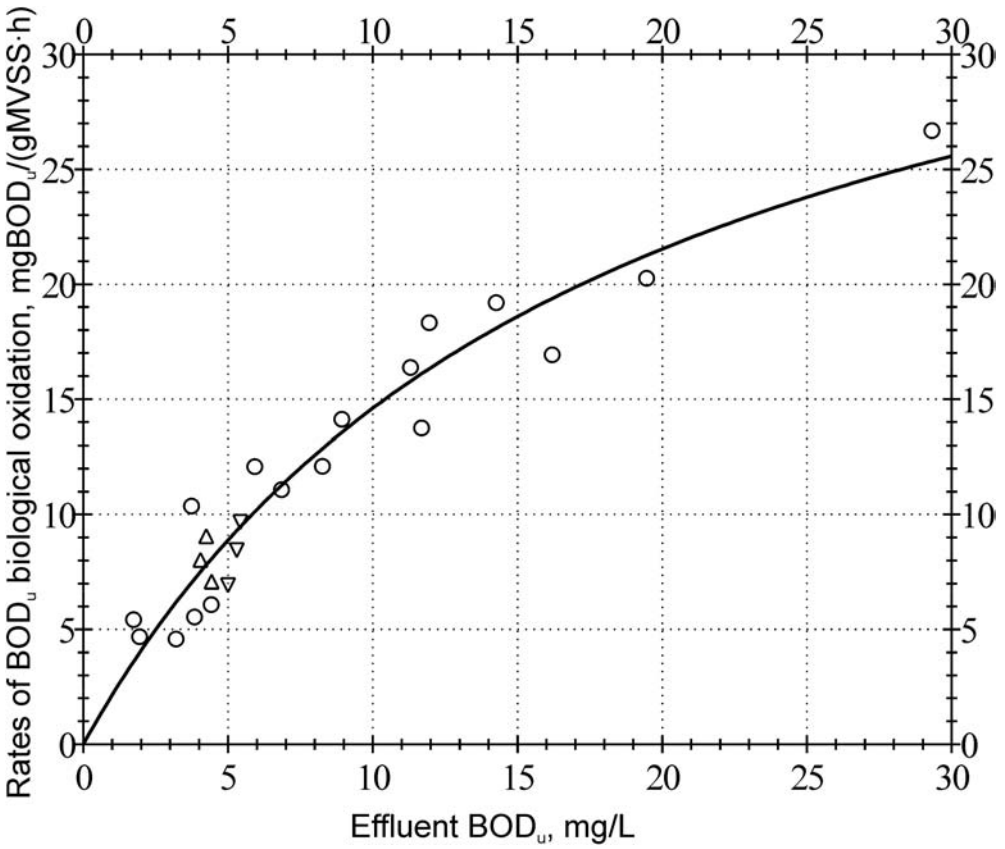

Fig. 9.8. Kinetic of the rates of BOD_u biological oxidation.

Therefore, research was continued at an HRT of 4.0 h and 6 h, SRT—13.8, 22, 32, and 43 days. The effluent nitrogen concentrations were: NH_4-N—0.08 to 0.2 mg/L; NO_2-N—0.06 to 0.017 mg/L; NO_3-N 0—0.35 mg/L; effectiveness of nitrification—99%; corresponding to efficiencies of denitrification—99%; TMLSS—5,800, 7,060 mg/L; COD—46.5 to 49.0 mg/L; BOD—4.0 to 6.2 mg/L; TSS—8.1 to 9.0 mg/L; ORP—80.0 to 150 mV.

The kinetic rates of biological oxidation are shown in Figs. 9.8 and 9.9. The plot in Fig. 9.8 developed for the rates of biological oxidation of organic meters based on BOD_u, at a TMLSS concentration of 5,900 mg/L, and dissolved oxygen concentration of 2 mg/L, indicates the following kinetic coefficients: $\rho_{max} = 40.9\ mgBOD_u/(gMVSS \cdot h)$; $K_m = 18.0$ mg/L in various types of aeration tank-clarifier (23). The plot in Fig. 9.9, developed for the rates of biological oxidation of organic matters based on COD at a TMLSS of 5,900 mg/L and dissolved oxygen concentration of 2 mg/L, indicates the following kinetic coefficients: $\rho_{max} = 90.7\ mgCOD/(gMVSS \cdot h)$; $K_m = 46.5$ mg/L; nonbiodegradable COD (S_H) = 32.3 mg/L.

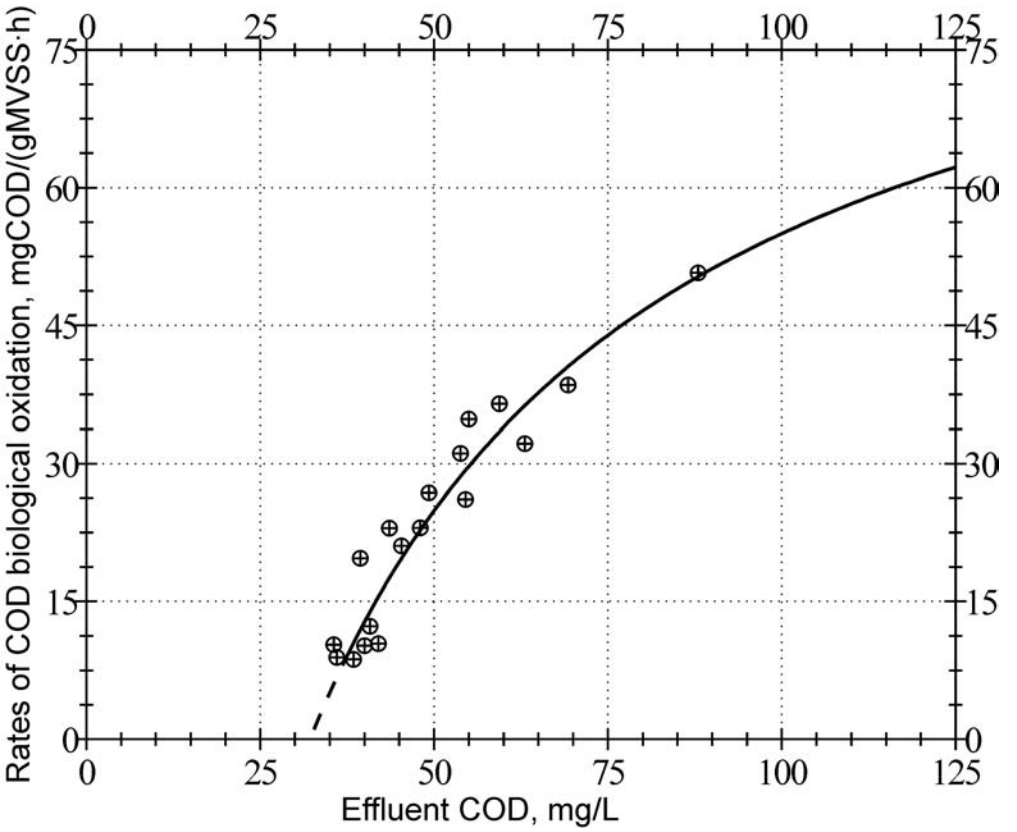

Fig. 9.9. Kinetic of the rates of COD biological oxidation.

The graphs show that this process is more efficient than conventional systems (23). Recirculation rates between aeration zones and clarification are in the range of 1,000% to 2,000% (23).

Research data on nitrification rates obtained from the operation of aeration tanks-clarifiers of the column and gallery-type at both pilot and full-scale levels are compared in Fig. 9.10 (Curve 1) with literature data of conventional biological systems as represented by Curves 2 and 3 (23–25). It must be asserted that the kinetic constants reported in Fig. 9.10 were obtained using nonlinear fitting of the experimental data to the Monod model. Curve 1 was obtained in the aeration tanks-clarifiers of the column and gallery-type at both pilot and full-scale level, with the kinetic coefficients: $\rho_{\max} = 20.8\,\text{mgN/(gMVSS} \cdot \text{h)}$; $K_m = 1.31\,\text{mgN/L}$. Curve 2 was obtained for industrial wastewater with high concentration of organic pollutants as well as nitrogen compounds from crystals lizin production, in conventional three stage nitrification-denitrification activated sludge-type facilities, with the kinetic coefficients: $\rho_{\max} = 19.4\,\text{mgN/(gMVSS h)}$; $K_m = 7.1\,\text{mg/L}$ (24). Curve 3 was obtained for

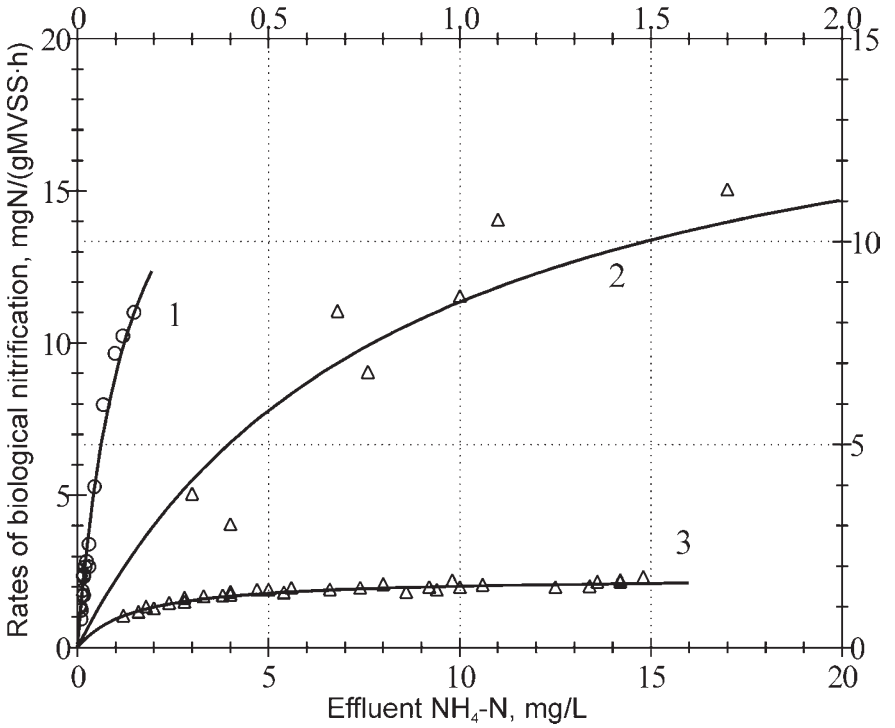

Fig. 9.10. Kinetic of the rates of biological nitrification.

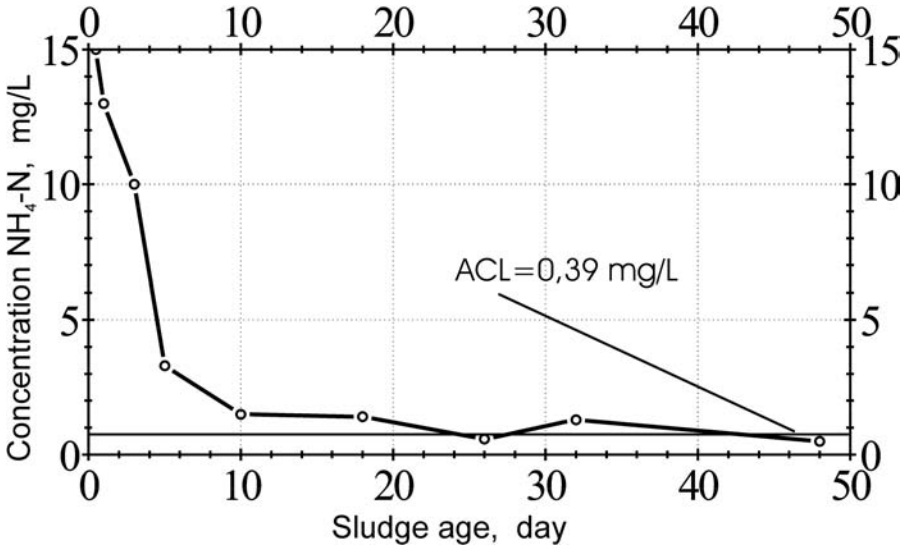

Fig. 9.11. The correlation between effluent NH_4-N concentration and solids residence time (SRT)—activated sludge age.

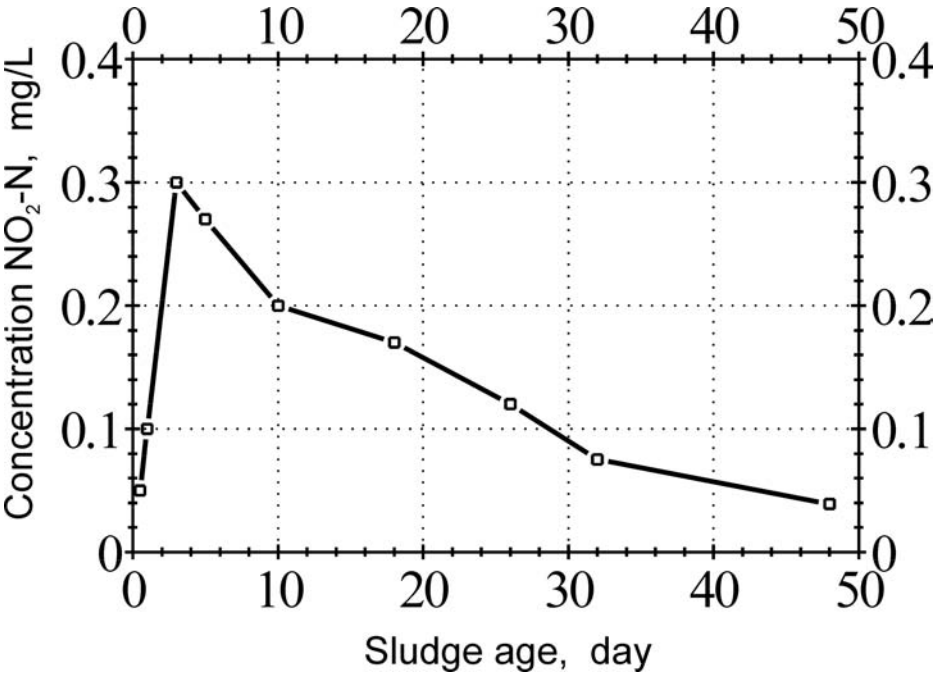

Fig. 9.12. The correlation between effluent NO_2-N concentration and solids residence time (SRT)—activated sludge age.

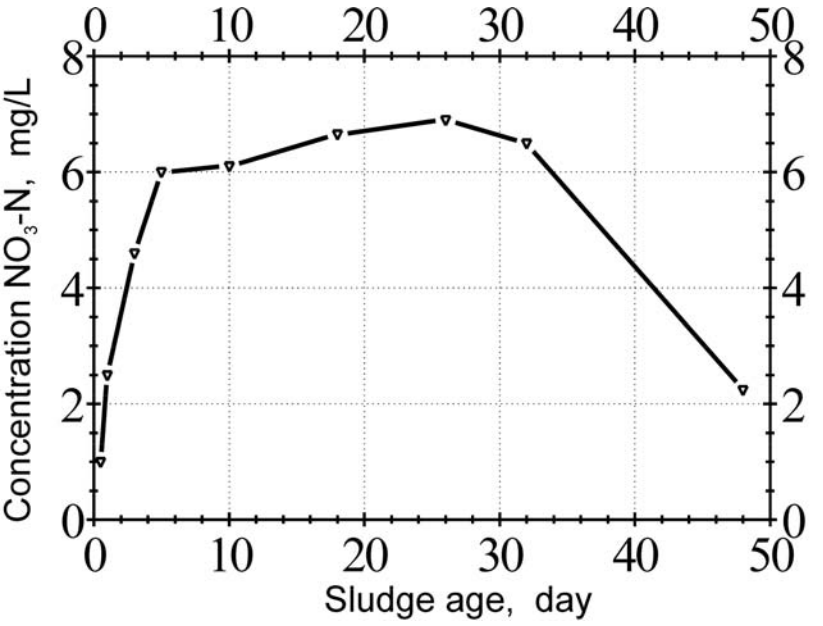

Fig. 9.13. The correlation between effluent NO_3-N concentration and solids residence time (SRT)—activated sludge age.

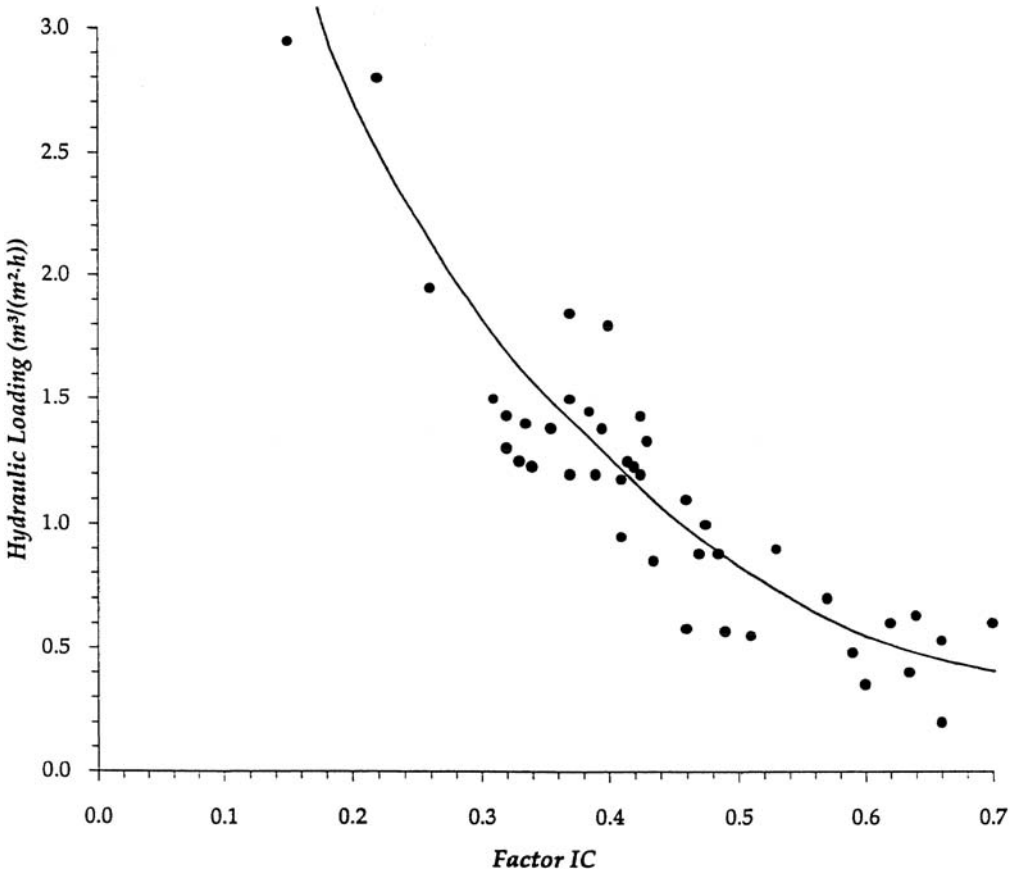

Fig. 9.14. The correlation between hydraulic loading on the separation surface of activated sludge and factor *IC*.

nitrification-denitrification of municipal wastewater in a conventional activated sludge system with the intermittent, with the kinetic coefficients: $\rho_{\max} = 2.3\,\text{mgN/(gMVSS} \cdot \text{h)}$; $K_m = 1.52\,\text{mg/L}$ (24).

Figure 9.10 with comparative data shows a higher nitrification rate in the aeration tanks-clarifiers. This phenomenon occurs because of the processes carried out in the pseudoliquified layers of activated sludge as an optimal reactor for biological nitrification and denitrification processes. It is also a result of improved contact and oxidation processes in the pseudoliquified layers of activated sludge and filtration of wastewater through flocs of activated sludge, thus enhancing substrate and oxygen transfer into the flocs. These data are correlated with other literature data (26–28). The figures show that this process is capable of additional biological nitrogen removal to the required effluent standards without the need to provide additional substrate.

The correlation between ammonia-nitrogen, nitrites, and nitrates removal efficiency and solid residence time (SRT), based on the research data obtained from the operation of aeration tanks-clarifiers of the column and gallery-type at both pilot and full-scale level and other literature data, is illustrated in Figs. 9.11–9.13 (23, 26–28). The graphs show that the increasing SRT improves ammonia-nitrogen removal efficiency as well as nitrites and nitrates efficiencies.

The correlation between hydraulic loading on the separation surface of activated sludge and factor *IC*, is presented in Fig. 9.14. The product of the sludge volume index (I) and TMLSS concentration (C) and calculated as follows:

$$IC = \frac{C \cdot I}{10^6} \tag{9}$$

where I is the sludge volume index, mL/g, C is the concentration of TMLSS, mg/L.

This research data was obtained in large-scale pilot plants with bioreactor volumes of 0.5 to 75.0 m^3, as well as actual aeration tank-clarifiers of column and gallery-types. The factor *IC* characterizes the settling characteristics of activated sludge and, therefore, efficiency of clarification. This graph shows that the hydraulic loading depends on the factor *IC* (i.e., TMLSS concentration) and that the stable existence of pseudoliquified layers is possible for the wide range of hydraulic loading of 0.75 to 3.0 $m^3/m^2/h$.

At MLSS concentrations in the range of 2,800 to 7,060 mg/L and organic loading of 0.196 to 0.92 mg COD/mg MLVSS, the specific yield of activated sludge varied from 0.21 to 0.39 mg MLVSS/mg BODu. Although these levels of sludge production correlate with those of extended aeration systems, the quantity of waste activated sludge produced is much less. Furthermore, complete nitrification and denitrification and phosphorus removal occur simultaneously.

4. DEVELOPMENT AND IMPLEMENTATION OF MODEL PILOT PLANT

4.1. System Capabilities and Need for Technology Refinement

Research and actual operating data indicate that the pseudoliquified activated sludge treatment is capable of providing secondary and tertiary treatment of municipal and/or industrial wastewater with a wide range of organic loading in single, two-stage, or multi-stage configurations, and produce a final effluent with BODu, total nitrogen, and TSS of approximately 5, 5 to 10, and 5 mg/L, respectively. Other process-related advantages include: absorption and oxidation processes rates that are up to twice as efficient as those of conventional systems owing to the favorable conditions for the microorganisms in the pseudoliquified layers of activated sludge and fixed film and the reduced return sludge cycle time. Operation of the system at TMLSS of 6,000 to 8,000 mg/L reduces the specific sludge loading and the rate of oxidation and amount of waste sludge. Furthermore, the process achieves much higher endogenous respiration rate, producing significantly less waste sludge.

Thus, the pseudoliquified activated sludge treatment is a viable and cost-effective technology for the treatment of municipal and industrial wastewaters. Based on over 10 years of operating experience in the Ukraine, Belarus, and Russia where the technology has been used

for municipal wastewater treatment in sizes ranging from 2,000 to 400, 000 m^3/day facilities, this technology is capable of achieving tertiary effluent quality with respect to removal of organics, nitrogen, and total suspended solids while achieving up to 80% reduction in land requirements and elimination of clarifiers. Operating data has demonstrated achievability of 95% nitrification and denitrification efficiency. The PA SB is capable of achieving effluent BOD, TSS, NH_3-N, and NO_3-N concentrations of 4 to 6, 8 to 10, 0.2 to 2.5 and 0 to 1 mg/L respectively.

With proper pretreatment, this technology can be used for treatment of industrial wastewaters to meet surface discharge criteria. However, because effluent phosphorus criteria for the plants that utilized this technology were not as stringent as North American and European standards of at least <1 mg/L, the technology was not designed to remove phosphorous. Furthermore, the technology did not achieve significant biological phosphorus uptake to realize effluent P of <1 mg/L. Accordingly, a new generation of this technology needed to be developed with chemical addition for P removal. The challenges with chemical addition at the long SRT in this system primarily result from accumulation of inorganic Fe and Al in the sludges, which might adversely influence biological activity. Thus a mobile pilot unit utilizing chemical addition for P removal was tested in Ukraine. This chemically enhanced pseudoliquified activated sludge bioreactor (CEPASB) was built and tested over a wide range of operational conditions to establish optimum design parameters and operating conditions.

4.2. Project Objectives

With the objective of successfully marketing this technology in other areas of the world, a mobile pilot plant was developed and tested jointly with Conestoga-Rovers and Associates with funding from the International Development Research Centre of Canada (IDRC).

The main technical objectives of the study were:

- To design a prototype of a column-type bioreactor combining biological nutrient removal with carbonaceous and organic matter removal processes. This prototype would then be used to establish wastewater treatability and process design criteria at specific sites to improve effluent quality.
- To manufacture and operate an advanced multistage biological and chemically enhanced biological treatment technology as a mobile pilot treatment unit to provide on-site treatment for small industrial and municipal wastewater treatment plants under local site conditions.
- To use the mobile unit to develop parameters for the optimization of existing treatment facilities. Ultimately, the optimization data would be used to design improvements to existing treatment facilities or to recommend the conceptual design for new facilities.
- To use the mobile unit to develop information on the benefits of various polymers to the treatment process.
- To use the mobile unit to provide an on-site training facility for operators of new and existing treatment facilities.

Refinement of the novel advanced wastewater treatment technology is a significant scientific advancement. Both short-term environmental benefits resulting from treatment trials and significant potential long-term environmental benefits from the implementation of the technology are emphasized. Additionally, it was surmised that substantial savings in

wastewater treatment plant upgrade costs may result. Realization of this project was made possible by the financial assistance from the Canadian International Development Agency (CIDA), administered by International Development Research Centre (IDRC). Additional project objectives included the creation of ecological management awareness in Ukraine and training of environmental experts.

4.3. Methodology

The technologies and plant designs used in this project generally produce effluents that meet world standards for treatment and are protected by patents (Nos. 7903522, 7907418, 604287, 1530573, 1638122, 1721028) (10–16).

The technology also evolved to incorporate the addition of specific polymers to increase biosolids concentrations and enhance treatment. This technology will be henceforth denoted as Chemically Enhanced Pseudoliquified Activated Sludge Bioreactor (CEPASB) technology.

Despite application at full-scale treatment plants in the Ukraine and Belarus, the PASB technology requires further refinement in the area of process control to achieve NH_3-N and TP concentrations of 0.39 and <1 mg/L, respectively. Optimization of the existing technology can be achieved through operational modifications such as biological solids residence time (SRT) control and chemical addition (i.e., alum or ferric for phosphorous removal). Furthermore, improved understanding of the technology and mathematical modeling of the various biochemical and physical processes along with the delineation of system specific parameters and biokinetic coefficients for various municipal and industrial applications is needed before the process can be marketed successfully in other areas of the world.

To achieve the objectives of this project, a modular pilot plant was built using the CEPASB technology to investigate on-site conditions at existing plant. Figures 9.15 and 9.16 demonstrate the layout and view of the mobile pilot plant. The mobile pilot unit was designed for the capacity of 25 m^3/day (3.81 gpm) and includes a complete biological treatment followed by tertiary filtration, as well as chemical reagent preparation and dosing chamber for phosphorus removal, and a waste-activated sludge measuring and storage tank.

Chemically enhanced pseudoliquified activated sludge bioreactor (CEPASB) technology is applied in a single deep column reactor incorporating biological oxidation of organics, nutrients removal (nitrogen and phosphorous), and clarification processes.

The biokinetic parameters are used for optimization of the existing technology for various municipal and industrial wastewater through operational modifications such as biological solids residence time (SRT), control and chemical addition for nitrogenous and phosphorus removal processes while minimizing biological sludge production.

4.4. Conceptual and Detailed Design of Mobile Pilot Plant

Conceptual and detailed designs of the mobile pilot unit were developed and include the following complete drafts: technological documentations; detailed drafts of installation and specification; architectural and planning details of the project including personnel and equipment area and plans for furniture and equipment installation; energy supply including electrical drawings of the project, control, and measuring equipment, plans for electrical

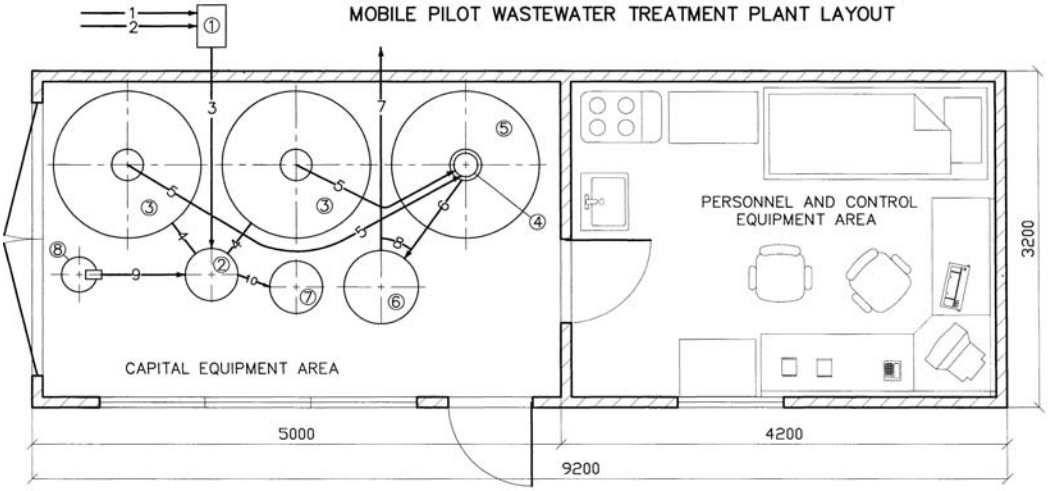

Fig. 9.15. Mobile pilot wastewater treatment plant layout.

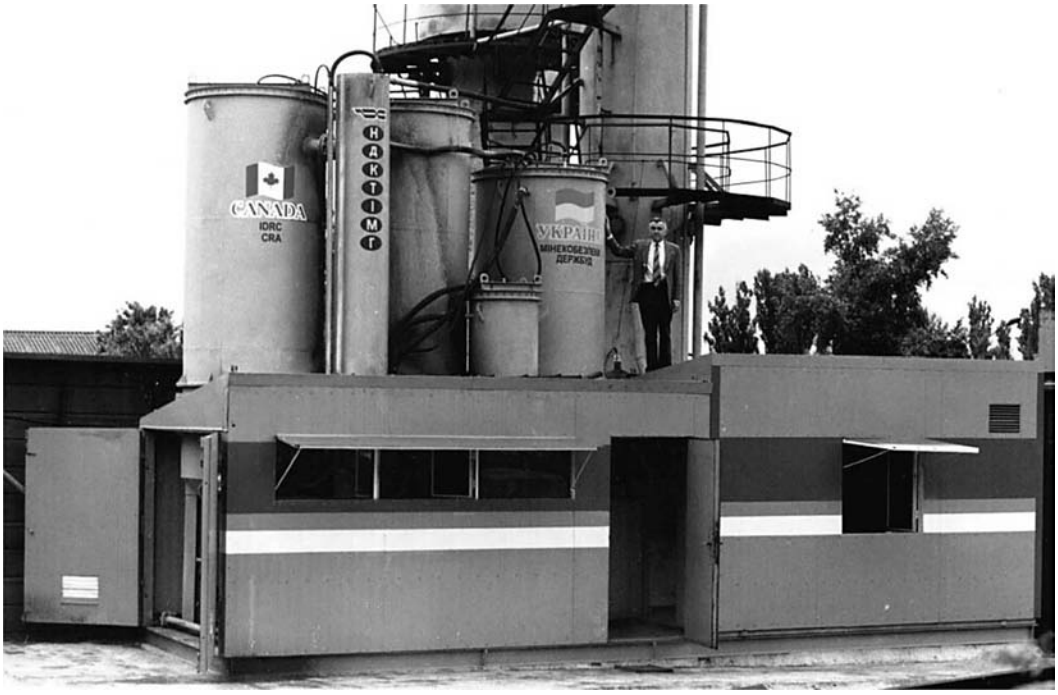

Fig. 9.16. Front view of the Mobile Pilot Wastewater Treatment Plant.

power control, master switch and power outlets; project facilities and personnel, control, and measuring equipment buildings.

4.5. Manufacturing, Installation, and Testing of the Mobile Pilot Plant

The main parts and facilities were manufactured and the mobile pilot unit constructed in accordance with the conceptual and detailed design developed.

The following parts were manufactured: facilities' columns, frames of the facilities, facilities within partitions, recirculation pipes, process pipes, installation of the facilities columns, and process piping of the mobile pilot unit. Control panels for the facilities and ancillary facilities (i.e., structures) to accommodate personnel, and measuring equipment were constructed at a manufacturing plant and transported on site on the trailer with a low platform.

The complete mobile pilot unit, including supporting infrastructure, was assembled on site at the Kiev Wastewater Treatment Plant. The entire treatment system was very compact and consisted of a 5×3.2 m facilities building and a 4.2×3.2 m personnel and equipment control building.

The mobile pilot plant was constructed in a manner so as to facilitate transportation to other testing sites. Hydraulic testing of the mobile pilot plant facilities (aeration column, clarifiers, bioreactor, tertiary filter, waste activated sludge storage tank, and chemical reagent) was carried out by using clean water to detect leaks.

Process pipes, and additional facilities and equipment, such as wastewater influent pipe with the flange gates from the pipe of municipal plant, influent wastewater and grit removal tank, pump for wastewater pumping into the pilot plant with the flange gates, primary settler with the pipes and flange gates, pipes, and wastewater measuring tank were put into operation during the hydraulic testing of the mobile pilot plant.

4.6. Development of Sampling and Monitoring Program

Details of the sampling and monitoring program were developed and provided to IDRC. The following highlights the important details related to this task. Jointly with the Conestoga-Rovers and Associates specialists, the detailed program of monitoring and sampling testing with all control parameters of the mobile pilot plant was developed. Process parameters to be analyzed were identified. The portable analytical equipment used for on-site testing of the pertinent water quality parameters and capital equipment were purchased for pilot plant testing in accordance with the project list and budget. The influent and effluent from the pseudoliquified activated sludge system as well as the effluent from the bioreactor were tested for total and soluble BOD, total and soluble COD, TKN, ammonia-nitrogen, nitrates-nitrogen, orthophosphates, total, and volatile suspended solids.

The mixed liquor and waste activated sludge were tested for total and volatile suspended solids. Dissolved oxygen concentration was monitored in the aeration tank and clarifier column tanks. The portable analytical apertures and capital equipment were shipped and received.

4.7. Testing of the Pilot Plant at Municipal Wastewater Facilities

The mobile pilot plant includes the following facilities (see Figs. 9.15 and 9.16):

- One aeration column—aeration tank: diameter—500 mm, height—5,100 mm, volume—1.0 m^3.
- Two clarifier column tanks: diameter—1,400 mm, height—4,427 mm, volume—11.8 m^3.
- One bioreactor column with the media: diameter—1,400 mm, height—4,427 mm, volume—5.23 m^3.
- One tertiary filter: diameter—700 mm, height of cylinder part—2,500 mm, height of cone part—227 mm, volume—0.77 m^3.
- One waste activated sludge measuring and storage tank: diameter—530 mm, height—720 mm, volume—0.12 m^3.
- Chemical reagent preparation and dosing tank: diameter—340 mm, height—1,050 mm, volume—0.08 m^3, with the mixing and a Metering Pump (gamma G/4b, ProMinent Fluid Controls, Inc.) for feeding ferric chloride solution, $FeCl_3$.

The mobile pilot plant was mounted with additional supporting facilities and equipment such as:

- A wastewater influent tank-grit removal: width—1,100 mm, length—1,100 mm, height—1,500 mm, volume—0.85 m^3.
- A centrifugal pump for wastewater pumping from wastewater influent tank grit removal into the pilot plant.
- A primary settler for suspended solid removal from wastewater: diameter—1,600 mm, height—2,000 mm, volume—4.0 m^3.
- A wastewater measuring and dosing tank into the aeration column of pilot plant: width—500 mm, length—1,000 mm, height—500 mm.
- A treated wastewater tank from pilot unit: diameter—1,000 mm, height—600 mm, volume—0.39 m^3.
- Process piping consisting of wastewater influent pipe with the well and flange gates from the pipe of municipal plant, treated wastewater discharge pipe, extra wastewater discharge pipe, waste-activated sludge discharge pipe, discharging pilot facilities pipes, influent air pipe with the gates and ball valves.

The mobile pilot plant was started with the addition of activated sludge. Three 3.0 m^3 tankers filled with activated sludge from the secondary settling tanks of Kiev wastewater treatment plant were added to the pilot plant. Mobile pilot plant testing was carried out in accordance with detailed testing, monitoring, and sampling program developed jointly with CRA.

The system was continuously operated from November 2000 to September 2001, spanning a period of 284 days. Mobile pilot plant testing were carried out in three stages with the hydraulic loading of 29.4 to 54.72 m^3/day (1.225, 1.7, 2.28 m^3/h).

Testing was carried out with actual Kiev treatment plant influent wastewater that comprised a mixture of municipal and industrial wastewater from the City of Kiev. Monitoring of process control parameters and chemical analysis of various effluents was performed during the mobile pilot plant testing.

Control of wastewater treatment processes in the pilot plant was carried out in a continuous manner with the following parameters measured daily: wastewater flow, air flow, waste activated sludge volume, grit removal sludge, primary settling sludge, sludge from

wastewater measuring tank, bioreactor sludge, tertiary filter sludge, and waste-activated sludge through the measuring tank.

Wastewater flow was measured using a measuring tank with a 90° weir. Wastewater flow through the 90° weir was calculated using the relation—$Q = 1.343 \times H^{2.47}$, where Q is the flow in m^3/S and H is the height in meters above the weir. The height of water level was measured by using a piezometer. A wastewater measuring tank was equipped with a flexible overflow device for discharge of extra wastewater and constant regulation of wastewater flow into the pilot plant.

Air flow in the aeration column of aeration tank-clarifier and bioreactor, was measured by flow meter, Model GT 1000, meter size 12, tube R-12M-20-5FT, float 12-RV-119, Brooks Instrument, USA.

Ferric chloride solution flow, $FeCl_3$, was measured and pumped by Metering Pump gamma G/4b, ProMinent Fluid Controls, Inc.

Process control parameters were analyzed by the following portable devices: wastewater quality parameters (i.e., COD, NH_3-N, NO_3-N, and PO_4-P) were analyzed using a Spectrophotometer DR/2010 Hach Company. Oxidation reduction potential (ORP) and pH analyses in the aeration tank-clarifier was monitored using an ORP/pH meter 500/mV 600 Series, hanna Instruments, pH-meter, HANNA Instruments. Dissolved oxygen concentration in the aeration tank was measured using a Dissolved Oxygen Meter, Model 2200 D/P91, AquaMetrix Inc., Canada. All other parameters were analyzed using standard analytical methods in accordance with the developed detailed testing, monitoring, and sampling program.

The results of mobile pilot plant testing data are illustrated in Tables. 9.2 to 9.7.

4.8. Detailed Analysis of Pilot Plant Testing Data

4.8.1. Commissioning Stage

The mobile pilot plant testing was conducted from November 22, 2000 to August 31, 2001, for a total of 284 days. Mobile pilot plant process testing was carried out in three stages with the hydraulic loadings of—29.4 to 54.72 m^3/day (1.225, 1.7, 2.28 m^3/h). The detailed analysis of the mobile pilot plant testing data was carried out. During the mobile pilot plant testing, investigation of a number of process control parameters and chemical analytical testing was undertaken.

Concentrations of various wastewater components at Kiev wastewater treatment plant are representative of combined municipal and industrial wastewater, and during the testing period were: oxidizable permanganate—30.0 to 450.0 mg/L; total BOD_5—40 to 460.0 mg/L; soluble BOD_5—26.8 to 151.0 mg/L; total COD—83.5 to 973.5 mg/L; soluble COD—32.5 to 695.2 mg/L; total suspended solids (TSS)—28.4 to 840.5 mg/L; total nitrogen, N—27.36 to 48.12 mg/L; ammonia-nitrogen, NH_4-N—13.3 to 36.72 mg/L; nitrites, $NO_2 - N$—0.00 to 4.88 mg/L; nitrates, NO_3-N—0.00 mg/L; total phosphorus, P—6.34 to 13.3 mg/L; orthophosphates, PO_4-P—0.98 to 11.08 mg/L.

Concentration of treated wastewater constituents after the aeration tank-clarifier of mobile pilot plant were: oxidizable permanganate—7.2 to 19.0 mg/L; total BOD_5—1.2 to 17.4 mg/L; soluble BOD_5—1.0 to 6.9 mg/L; total COD—26.8 to 95.6 mg/L; soluble COD—24.4 to

Table 9.2
Total nitrogen removal in mobile pilot plant

		Total nitrogen, N, mg/L		NH_4-N, mg/L			Nitrites, NO_2-N, mg/L			Nitrates. NO_3-N, mg/L	
Date	t, °C	Primary effluent (soluble)	Aeration tank (sol)	Primary effluent (sol)	Aeration tank (sol)	Bioreactor (sol)	Primary effluent (sol)	Aeration tank (sol)	Bioreactor (sol)	Aeration tank (sol)	Bioreactor (sol)
08/08/01	28	27.36	2.61	14.8	1.5	–	0.02	0.06	-	1.05	-
10/08/01	28.8	–	–	13.3	2.18	–	0.18	0.06	-	1.78	-
15/08/01	26.4	48.12	3.63	23.3	0.62	–	0.05	0.03	-	1.51	
16/08/01	23	37.79	3.10	–	–	–	–	–	-	-	-
17/08/01	24.8	42	3.63	34.5	2.1	–	0	0	-	0	-
20/08/01	28	36.44	3.12		–	–	–	–	-	-	-
22/08/01	29	36.44	3.90	30.7	0.27	0.19	0	0	0.03	3.6	2.39
28/08/01	25	32.5	3.91	–	–	–	–	–	-	-	-
29/08/01	24	32.5	2.73	28.3	2.65	1.64	0.09	0.06	0.03	1.27	2.26
30/08/01	20	30.53	2.73	19.8	0.93	0.62	0.1	0.006	0.006	2.9	2.8

Table 9.3
Total phosphorus removal in mobile pilot plant

Date	t, °C	Ferric solution concentration, $FeCl_3$, g/L (%)	Ferric solution Flow rate, L/h	Concentration, $FeCl_3$/Fe, mg/L	Orthophosphates, PO_4-P, mg/L		Total phosphorus, P, mg/L	
					Grit removal effluent (soluble)	Aeration tank effluent (soluble)	Grit removal effluent (soluble)	Aeration tank effluent (soluble)
17/07/01	28	25 (2.5)	1.44	16/3.3	3.17	2.32	–	–
18/07/01	28	25 (2.5)	1.44	16/3.3	–	–	–	–
19/07/01	26.5	25 (2.5)	1.44	16/3.3	–	–	–	–
20/07/01	28	25 (2.5)	1.44	16/3.3	–	–	–	–
21/07/01	28	25 (2.5)	1.44	16/3.3	–	–	–	–
22/07/01	29	25 (2.5)	1.44	16/3.3	–	–	–	–
23/07/01	27	50 (5)	1.44	32/6.6	–	–	–	–
24/07/01	28	50 (5)	1.44	32/6.6	–	–	–	–
25/07/01	28	50 (5)	1.44	32/6.6	3.8	1.96	12	3.3
26/07/01	28.5	50 (5)	1.5	32.9/6.8	–	–	–	–
27/07/01	28	50 (5)	1.5	32.9/6.8	4.9	1.57	11.2	2.35
28/07/01	28	50 (5)	1.5	32.9/6.8	–	–	–	–
29/07/01	28	50 (5)	1.5	32.9/6.8	–	–	–	–
30/07/01	28	50 (5)	1.5	32.9/6.8	–	–	–	–
31/07/01	29.7	50 (5)	1.5	32.9/6.8	–	–	–	–
01/08/01	27.5	50 (5)	1.5	32.9/6.8	3.68	1.32		–
02/08/01	27.5	50 (5)	1.5	32.9/6.8	–	–	7.31	2.1
03/08/01	25	50 (5)	1.5	32.9/6.8	2.84	0.91	–	–
08/08/01	28	0	0	0	3.69	2.28	–	–
10/08/01	28.8	0	0	0	3.04	2.7	–	–
15/08/01	26.4	0	0	0	3.53	2.9	–	–
17/08/01	24.8	0	0	0	7.67	3.1	–	–
21/08/01	28	62 (6.2)	3.18	86.5/18	–	–	–	–
22/08/01	28	62 (6.2)	3.18	86.5/18	2.12	0.5	6.34	0.89
23/08/01	27	52 (5.2)	3.18	72.5/15	4.1	0.5	9.8	0.79
24/08/01	27	52 (5.2)	3.18	72.5/15	4.6	0.5	11.0	0.75
25/08/01	26	52 (5.2)	3.18	72.5/15	4.0	0.4	10.2	0.63
26/08/01	26	58 (5.8)	3.18	80.9/16.7	5.2	0.5	13.3	0.76
27/08/01	26	100 (10)	3.18	139.5/28.9	4.7	0.5	11.2	0.75
28/08/01	26	80 (8)	3.18	111.6/23.1	4.2	0.3	7.76	0.58
29/08/01	25	55 (5.5)	3.18	76.7/15.8	2.41	0.29	6.84	0.56
30/08/01	20	55 (5.5)	3.18	76.7/15.8	3.46	0.42	8.7	0.77

67.7 mg/L; TSS—1.8 to 21.2 mg/L; total nitrogen, N—2.61 to 3.91 mg/L; ammonia-nitrogen, NH_4-N—0.27 to 2.65 mg/L; nitrites, NO_2-N—0.00 to 0.06 mg/L; nitrates, NO_3-N—0.00 to 2.9 mg/L; total phosphorus, P—0.56 to 0.89 mg/L; orthophosphates, PO_4-P—0.29 to 0.5 mg/L.

Wastewater hydraulic retention time (HRT) in the aeration tank-clarifier was in the range 4.8 to 8.9 hours. During this wastewater retention time, complete biological nitrification and denitrification processes took place in the single aeration tank-clarifier. This system is very economical, compared to a conventional aeration tank with separate reactors for nitrification

and denitrification, with HRTs of 10 to 16 hours in the bioreactors, and plus additional 4 to 6 hours for settling in the secondary settler.

Total mixed liquor suspended solids (MLSS) concentrations during testing period in the aeration tank-clarifier were in the range of 4,100 to 8,400 mg/L, whereas volatile suspended solids (MLVSS) were 3,063 to 6,426 mg/L. By comparison with separate aeration tank and secondary settling usually operating with TMLSS concentrations in the range of 2,000 to 2,500 mg/L, solids concentration in this system are 2 to 2.8 times higher than conventional systems, thus affecting the excellent removal efficiencies within much shorter HRTs.

Dissolved oxygen concentration in the aeration tank-clarifier during testing period was in the range of 1.52 to 10.5 mg/L. Airflow rates to the aeration tank-clarifier were in the range 6.35 to 35.65 m^3/h, corresponding to specific flow rate of air per unit flow of wastewater (air/ww), in the range of 2.8 to 15.6 m^3/m^3.

4.8.2. Optimization of Operational Parameters for Activated Sludge—Clarifier System

4.8.2.1. Optimization of Loading Rate

The hydraulic loading rates were in the range—29.4 to 54.72 m^3/day (1.225, 1.7, 2.28 m^3/h), during testing of the mobile pilot plant. Table. 9.2 and 9.3 present representative analytical data related to N and P removal in the CEPASB at the maximum hydraulic loading rate.

The organic loading rates based on BOD_5, in accordance with the hydraulic loading rates wastewater concentration and volatile suspended solids (VSS), of activated sludge at the aforementioned loading were in the following ranges: total BOD_5 0.026 to 0.36 mg BOD_5/mg VSS/day; soluble BOD_5 0.014 to 0.098 mg BOD_5/mg VSS/day.

Table 9.4 presents the food-to-microorganisms ratio and sludge yields for the CEPASB at the three hydraulic loadings investigated in the study. The CEPASB operated at F/M ratios of 0.06 to 0.46 mg BOD_5/mgVSS/day over the entire testing period with average F/M ratios for the 20.4, 40.8, and 54.7 m^3/day of 0.093, 0.099, and 0.14 mg BOD_5/mgVSS/day, respectively. These biomass specific organic loading rates fall in the range between the upper end for extended aeration of 0.1 mg BOD_5/mgVSS/day and the lower end of conventional plants of 0.2 mg BOD_5/mgVSS/day. It is interesting to note that despite the very short HRT of 4 to 6 hours, biomass specific organic loading rates are typical of extended aeration systems as a result of the high MLVSS concentrations. The low sludge yields ranging from 0.15 to 0.27 g TMSS/g BOD_5 and 0.23 to 0.3 g TMSS/g COD are noteworthy. By comparison with conventional activated sludge plants with typical yields of 0.5 g VSS/g BOD_5 and volatile suspended solids fractions of 0.67 (i.e., corresponding to an overall yield of 0.75 g TMSS/g BOD_5) this system produces 64% to 80% less sludge. Furthermore, by comparison with activated sludge biological nitrogen removal systems with a typical yield of 0.4 g VSS/g COD and a volatile suspended solids fraction of 0.65 (i.e., equivalent to an overall yield of 0.62 g MSS/g COD), this technology produces 52% to 63% less sludges.

Table 9.4
Variation of biomass-specific organic loadings and biomass yields

Hydraulic Lodaing Rate	Organic loading								Sludge Yield			
	F:M (BOD based)				F:M (COD based)							
	BOD_5		$SBOD_5$		COD		SCOD		BOD Based		COD Based	
	mg BOD_5/mg VSS day		mg $SBOD_5$/mg VSS day		mg COD/mg VSS day		mg SCOD/mg VSS day		g TMSS/g BOD_5		g TMSS/g COD	
m^3/h	Range	Average	Range	Average	Range	Average	Range	Average	Range	Average	Range	Average
1.225	0.065–0.122	0.093	0.02–0.047	0.031	0.08–0.3	0.17	0.03–0.09	0.047	0.06–0.26	0.153	0.15–0.34	0.227
1.7	0.06–0.36	0.099	0.014–0.098	0.038	0.1–1.143	0.211	0.024–0.164	0.073	0.04–0.62	0.266	0.1–0.5	0.296
2.28	0.057–0.455	0.14	0.028–0.071	0.05	0.085–1.11	0.235	0.04–0.12	0.054	0.03–0.65	0.174	0.07–0.68	0.257

4.8.2.2. Impact of Biological Solids Residence Time (SRT)—Sludge Age

Biological solid residence time (SRT) or sludge wastage rates in the system were controlled constantly during testing of the mobile pilot plant. The volume of waste activated sludge in the aeration tank was measured every day using the 120-L waste activated sludge tank.

Biological SRT was estimated based on the quantity of waste activated sludge from the aeration tank whereas the real SRT factored in the amount of solids leaving in the system effluent and consequently real SRTs were below calculated SRTs.

The variation of the biological SRT of mobile pilot plant during the testing period were in the range: SRT (calculated)—15.1 to 64.1 days; SRT (actual)—12.0 to 36.4 days.

Minimum SRTs were investigated at the highest hydraulic loading rate of 54.7 m^3/day. At the relatively short SRTs in the range of 12 to 17 days and the maximum hydraulic loading rate of 2.28 m^3/h (54.72 m^3/day) excellent removal efficiencies were achieved as apparent from Table. 9.1 and 9.2. It should be asserted that the combination of highest hydraulic loading with short SRTs represents the worst case scenario with respect to treatment efficiency, yet the system performed remarkably well.

4.8.3. Nutrient Removal

The mobile pilot plant constantly achieved high nutrient removal efficiency. Various forms of nitrogen (i.e., ammonia-nitrogen (NH_4-N)) nitrites ($NO_2 - N$), nitrates (NO_3-N), and total nitrogen (N) were analyzed. Following below is a summary of the various N species in the influent and effluent of the system: pilot plant influent: total nitrogen, N—27.36 to 48.12 mg/L; ammonia-nitrogen NH_4-N—13.3 to 36.72 mg/L; nitrites, NO_2-N—4.88 to 0.00 mg/L; nitrates NO_3-N—0.00 mg/L; treated wastewater (aeration tank effluent)—total nitrogen, N—2.61 to 3.91 mg/L; ammonia-nitrogen, NH_4-N—0.27 to 2.65 mg/L; nitrites, NO_2-N—0.00 to 0.06 mg/L; and nitrates, NO_3-N—0.00 to 2.9 mg/L.

Nitrogen removal rates with respect to: total nitrogen, N, ammonia-nitrogen, NH_4-N, nitrites, NO_2-N, and nitrates, NO_3-N are illustrated in Table 9.2.

As apparent from Table 9.2, an optimal operational range was achieved at a hydraulic rate of 2.28 m^3/h (54.72 m^3/day) (from August 08, 2001 to August 30, 2001), during which the following performance was achieved. Pilot plant influent: total nitrogen, N—27.36 to 48.12 mg/L; ammonia-nitrogen, NH_4-N—13.3 to 34.5 mg/L; nitrites, NO_2-N—0.18 to 0.00 mg/L; nitrates, NO_3-N—0.00 mg/L; treated wastewater (aeration tank effluent): total nitrogen, N—2.61 to 3.91 mg/L; ammonia-nitrogen, NH_4-N—0.27 to 2.1 mg/L; nitrites, NO_2-N—0.00 to 0.06 mg/L; nitrates, NO_3-N—0.00 to 2.9 mg/L.

It must be emphasized that nitrogen removal efficiencies achieved in this system, as well as effluent quality nitrogen parameters meet the criteria for discharge to rivers in Ukraine.

Based on the excellent nitrogen removal efficiencies in conjunction with organics and suspended solids removal, the following optimal process parameters derived from the pilot plant are recommended for design and implementation of the technology: hydraulic loading 2.28 m^3/h (54.72 m^3/day); wastewater retention time (contact with activated sludge) 4.78 hours; biological solid residence time (SRT)12 to 17 days; average organic loading rate based on total BOD_5 0.14 mg BOD_5/mg VSS d; average organic loading rate based on soluble BOD_5 0.05 mg BOD_5/mg VSS/day; average organic loading rate based on total COD

0.235 mg COD/mg VSS/day; average organic loading rate based on soluble COD 0.054 mg COD/mg VSS/day; activated sludge yield based on total COD 0.07 to 0.68 g TMSS/g COD, average rate 0.257 g TMSS/g COD; activated sludge yield based on total BOD_5 0.03 to 0.65 g TMSS/g, BOD_5 average rate—0.174 g TMSS/g BOD_5.

4.8.4. Phosphorus Removal

Phosphorus removal was constantly monitored during the testing period of the mobile unit by analysis of orthophosphate, PO_4-P, and total phosphorus P concentrations.

A summary of the removal of orthophosphates (PO_4-P) and total phosphorus (P) for the three hydraulic loadings investigated in this study is shown in Table 9.5. It should be noted that during the 29.4 and 40.8 m^3/day hydraulic loadings, chemical addition for P removal was not undertaken. During the 29.4 m^3/day, influent orthophosphates (as P) in the range of 3.9 to 9.9 mg/L were reduced by 45% to 82% to 0.1 to 3.4 mg/L. Similarly during the 40.8 m^3/day loading, influent PO_4-P of 2.1 to 14.9 mg/L were reduced to 0.2 to 3.4 mg/L. Accordingly during these two conditions, phosphorus removals of 1.9 to 11.5 mg/L were achieved biologically without chemical addition. On average 3 to 4 mg/L of PO_4-P were removed biologically. Because the raw wastewater BOD was mostly in the 100 to 150 mg/L range, approximately 1 to 1.5 mg/L of P would be consumed for bacterial synthesis. Although this data may suggest enhanced P removal in such systems, this was not substantiated by analysis of the waste sludge.

As apparent from the above data, orthophosphate removal efficiencies achieved with biological assimilation were unstable, despite occasionally high rates in the range of 84.7% to 95.8%. Nonetheless, the system could not achieve effluent TP of <1 mg/L, which is a typical effluent standard in North America.

To achieve more stable phosphorous removal efficiencies both with respect to orthophosphate, PO_4-P, and total phosphorus, the testing was carried out with addition of ferric chloride, $FeCl_3$, solution to the aeration tank from July 17, 2001 to August 30, 2001. The mobile pilot plant operation with ferric chloride addition was carried out in two periods from July 17, 2001 to August 03, 2001 and from August 21, 2001 to August 30, 2001.

Table 9.5
Detailed biological and phosphorous removal in the mobile pilot unit

		Influent phosphorus		Effluent phosphorus		Percent removal	
Hydraulic Lodaing Rate m^3/h	Ferric Chloride Dose mg/L as Fe	PO_4-P mg/L	TP mg/L	PO_4-P mg/L	TP mg/L	PO_4-P %	TP %
1.225	0	3.88–9.93	–	0.1–3.4	–	45.0–82.0	–
1.7	0	2.12–14.87	–	0.2–3.4	–	0–90.5	–
2.28	0	0.98–4.08	–	0.1–3.4	–	27.0–95.8	–
2.28	0	0.98–4.67	–	0.42–3.73	–	0.0–84.7	–
2.28	3.3–6.8	2.84–4.9	7.31–12.0	0.91–2.32	2.1–3.3	27.0–68.0	71.0–79.0
2.28	15.0–28.9	2.12–5.2	6.34–13.3	0.3–0.5	0.56–0.89	76.4–93.1	86.0–94.4

Ferric chloride solution was prepared in a 0.08 m^3 tank equipped with high-speed mixing and metering pump for pumping solution into the pilot plant. Ferric chloride solution was pumped in the aeration column of aeration tank of pilot plant.

The optimum ferric chloride solution concentration, flow rate, and dose of ferric chloride for complete removal of orthophosphate, PO_4-P, and total phosphorus, P was investigated during the first period—from July 17, 2001 to August 03, 2001.

As apparent from Table 9.5, ferric addition at dosages in the range of 3.3 to 6.8 mg/L as Fe failed to reduce effluent concentrations TP, below 1 mg/L. However, ferric chloride doses corresponding to 15 to 29 mg/L as Fe readily achieved the TP criteria of <1 mg/L. To achieve P removals of $>90\%$, molar ratios of Fe:P of 2:1 should be maintained and accordingly a mass ratio of 2.9 mg Fe/mg P is needed. This compares very well with the optimum Fe doses of 15 to 18 mg/L corresponding to Fe : PO_4-P mass ratios of 2.9:1 to 3.5:1. This suggests that enhanced biological P removal did not occur in the pseudoliquified activated sludge bioreactor.

During this testing phase with chemical addition, the system achieved the following performance.

The pilot plant wastewater influent: orthophosphate, PO_4-P 2.12 to 5.2 mg/L; total phosphorus, P 6.34 to 13.3 mg/L; treated wastewater aeration tank effluent (with addition of ferric chloride, $FeCl_3$,); orthophosphate, PO_4-P 0.3 to 0.5 mg/L; and total phosphorus, P 0.56 to 0.89 mg/L.

Effectiveness of phosphate removal was: orthophosphate, PO_4-P, in the range—76.4% to 93.1%; total phosphorus, P, in the range 86.0% to 94.4%.

Thus, the optimum dose of ferric chloride, was in the range of 72.5 or 15.0 mg/L (Fe) to 80.9 or 16.7 mg/L (Fe), at which the standard total phosphorus concentrations of <1 mg/L were achieved with the system achieving the following effluent concentrations: orthophosphate, PO_4-P 0.29 to 0.5 mg/L; total phosphorus, P 0.56 to 0.76 mg/L.

Furthermore, it must be emphasized that at full-scale wastewater treatment plant operation, the optimum dose of ferric chloride will be much less than that obtained during this experimental pilot plant testing owing to accumulation of ferric chloride into suspended layers of activated sludge.

Solids concentration in the aeration tank was used in conjunction with 30 min settling data to estimate sludge volume index (SVI) and assess settleability, during the mobile pilot plant testing period. Sludge volume indices (SVIs), were mostly in the range of 123 to 50 mL/g, without ferric chloride ($FeCl_3$), addition while with ferric, SVIs dropped precipitously to 40 to 50 mL/g, reflecting excellent settleability.

4.8.5. Achievability of Effluent Criteria

Based on the results of the pilot testing, the optimum process parameters are: hydraulic loading of 2.28 m^3/h (54.72 m^3/day); wastewater HRT (contact with activated sludge) of 4.78 hours; and biological SRT of 12 to 17 days, with a corresponding effluent quality of: total BOD_5 1.2 to 5.6 mg/L. Soluble BOD_5—1.0 to 4.9 mg/L; total COD—37.2 to 66.6 mg/L; soluble COD—24.4 to 65.0 mg/L; total suspended solids—1.8 to 15.0 mg/L; total nitrogen,

N—2.61 to 3.91 mg/L; ammonia-nitrogen, NH_4-N—0.27 to 2.1 mg/L; nitrites, NO_2-N 0.00—0.06 mg/L; nitrates, NO_3-N 0.00—2.9 mg/L; orthophosphate, PO_4-P—0.29 to 0.5 mg/L; and total phosphorus, P—0.56 to 0.76 mg/L.

As mentioned above, the standard total phosphorous criterion of less than 1 mg/L was achieved with an optimum dose of ferric chloride, in the range of 72.5 or 15.0 mg/L (Fe) to 80.9 or 16.7 mg/L (Fe).

Microbiological investigations of activated sludge were carried out during the pilot plant testing. The activated sludge had a swamp odor and a brown-gray color. Microscopic examination of the activated sludge revealed the presence of low-class organisms including bacteria, fungi, algae, and simple organisms such as flagellate, sarcodic, vibratile infusorium, predation infusorium, nematode, and rotatoria. Vibratile suctorial infusorium, nematode, and rotatoria are the representatives of third trophic level of the reservoir and river. The presence of third trophic level of microorganisms in the activated sludge is indicative of a high level of wastewater treatment and high quality of treated wastewater.

The pathogenic microorganism removal was investigated during the testing period of the pilot plant by bacteriological analysis of pathogen bacteria concentration. Following below are the overall ranges of total coliform counts per liter of wastewater in the influent and treated effluent: wastewater influent (grit removal tank effluent): 43×10^4 to 4×10^7; treated effluent from aeration tank-clarifier: 37×10^2 to 1.9×10^4; and treated effluent from tertiary bioreactor: 5×10^2 to 1×10^4. The results of bacteriological analysis as well as microbiological investigations of activated sludge are indicative of a high oxidative state in the system and high quality of treated wastewater.

Towards the end of operation at the optimum loading condition of an HRT of 4.8 hours and an SRT of 15 days, development of mass daphnia was obvious. This development was evident not only in the treated wastewater zone but also in all volume of activated sludge mixture, as well as the clarification and aeration zones indicative of high oxidative state and purity of the wastewater.

4.8.6. Tertiary Biological Treatment

4.8.6.1. System Description and Operational Parameters

Biologically treated wastewater from aeration tank-clarifier was further processed in a tertiary bioreactor. The bioreactor was a column with media: diameter—1,400 mm, height—4,427 mm, and volume—5.23 m^3.

The bioreactor was used for tertiary biological treatment of wastewater for additional removal of: BOD, COD, total suspended solids, nitrogen, and phosphorus by immobilized microorganisms owing to wastewater filtration through the media with microorganisms. Bioreactor media were made from braided kapron fiber that looks like a "WII" (CILIA), which are extended on a frame. The bacteria, fungi, algae, and simple organisms grew on the media surface as a biological film. A perforated pipe in the central column of the bioreactor served as a diffuser for wastewater aeration and dissolved oxygen saturation. Biological tertiary treated wastewater was collected by a circular pipe and came through a cone funnel into the filter.

The bioreactor was tested at hydraulic loading rates of: 1.225 (29.4 m^3/day), 1.7 (40.8 m^3/day), and 2.28 m^3/h (54.72 m^3/day). Wastewater retention time in the bioreactor

were: 4.2, 3, and 2.3 hours and the hydraulic loading rates based on cross-sectional area of the bioreactor (1.13 m^2) were, respectively: 1.08 (0.3 mm/s), 1.5 (0.42 mm/s), and 2.02 $m^3/(m^2.h)$ (0.56 mm/s).

4.8.6.2. General Performance and Effluent Quality

At the following operational conditions (i.e., hydraulic loading rate of 1.225 m^3/h (29.4 m^3/day) HRT of 4.2 hours, and surface loading of 1.08 $m^3/m^2.h$ and the following effluent quality was achieved: total BOD_5—2.3 to 3.5 mg/L; soluble BOD_5—1.4 to 3.0 mg/L; total COD—41.5 70.3 mg/L; soluble COD—40.2 to 69.8 mg/L; total Suspended Solids—2.2 to 8.4 mg/L; ammonia-nitrogen, NH_4-N—0.5 to 2.1 mg/L; nitrites, NO_2-N—0.05 to 0.07 mg/L; nitrates, NO_3-N r—0.00 to 1.13 mg/L; orthophosphates, PO_4-P, (without addition of ferric chloride solution, $FeCl_3$)—4.25 to 9.6 mg/L.

At the hydraulic loading of 1.7 m^3/h (40.8 m^3/day) HRT of 3.0 hours and surface loading of 1.5 $m^3/m^2.h$, the following effluent quality was achieved: total BOD_5—2.0 to 9.32 mg/L; soluble BOD_5—1.9 to 8.12 mg/L; total COD—31.0 to 61.1 mg/L; soluble COD—30.3 to 59.3 mg/L; total suspended solids—0.2 to 15.0 mg/L; ammonia-nitrogen, NH_4-N—0.03 to 2.18 mg/L; nitrites, NO_2-N—0.00 to 0.12 mg/L; nitrates, NO_3-N—0.00 to 2.47 mg/L; orthophosphates, PO_4-P, (without addition of ferric chloride solution, $FeCl_3$)—0.05 to 6.93 mg/L.

At the hydraulic loading of 2.28 m^3/h (54.7 m^3/day), HRT of 2.3 hours, and surface loading of 2.02 $m^3/m^2/h$, the following effluent quality was achieved: total BOD_5—2.7 to 6.8 mg/L, soluble BOD_5—1.9 to 6.2 mg/L, total COD—40.8 to 68.2 mg/L, soluble COD—39.6 to 67.6 mg/L, total suspended solids—0.2 to 15.0 mg/L, ammonia-nitrogen, NH_4-N—0.19 to 1.64 mg/L, nitrites, NO_2-N—0.006 to 0.03 mg/L, nitrates, NO_3-N—0.00 to 2.39 mg/L, orthophosphates, PO_4-P, (with addition of ferric chloride solution, $FeCl_3$)—0.42 to 0.59 mg/L.

As discerned from the aforementioned data and by comparison with the aeration tank effluent, the effluent wastewater quality improved marginally across the bioreactor only at the hydraulic loading rate of 29.4 m^3/day. However, after increasing the hydraulic loading to 1.7 (40.8 m^3/day), and 2.28 m^3/h (54.72 m^3/day), effluent quality did not improve. This can be explained in terms of increasing biomass thickness owing to suspended solids accumulation on surface media and increasing shearing of suspended solids resulting in deteriorating quality effluent after bioreactor at high loadings.

The standard effluent phosphorus criterion of less than 1 mg/L was achieved after the bioreactor with ferric chloride optimum doses, in the range 72.5 or 15.0 mg/L (Fe) to 80.9 or 16.7 mg/L (Fe).

4.8.7. Process Economics

Cost-effectiveness of the CEPASB is determined in terms of reduction of both capital and operation costs, owing to lower energy consumption, process parameters optimization and lower sludge production, as well as reduction of land requirements by up to 80%. Economics of biological wastewater treatment in the pilot plant are determined on the basis of specific energy consumption for wastewater pumping and supply for aeration in the aeration tank and bioreactor.

Energy consumption for wastewater pumping in the plant was determined as follows.

Power used for wastewater pumping in the plant is 1.1 kilowatt (kW). The pump wastewater output is 8.0 m^3/h, and accordingly, specific energy consumption on wastewater pumping in the plant is—$1.1/8.0 = 0.13\ kW/m^3$. Therefore, energy consumption for wastewater pumping in the plant at the hydraulic loading of 1.225 m^3/h (1.7 and 2.28 m^3/h), 29.4, 40.8, and 57.7 m^3/day are 0.16 and 0.221 kWh, respectively. Energy consumption for wastewater pumping in the plant for the hydraulic loading of 2.28 m^3/h (54.72 m^3/day) $- 2.28 \times 0.13 =$ 0.296 kWh.

Specific energy consumption for air delivery in the plant is $0.035\ kW/m^3$. Energy consumption for air delivery in the plant at the different airflow rates into the aeration tank in the range of 6.35 to 33.7 m^3/h is this 0.222 to 1.18 kWh, respectively. Energy consumption for air delivery to the bioreactor is thus 0.101 kWh. Therefore the total energy consumption for wastewater pumps and air delivery to the CEPASB at the hydraulic loading 2.28 m^3/h (54.72 m^3/day), based on the average aeration energy of 0.7 kWh is 1.0 kWh, which translates to $0.44\ kWh/m^3$ of flow. This value is at the lower end of the range for biological nutrient removal plants in Germany of 0.4 to $0.8\ kWh/m^3$ (31).

Capital cost-effectiveness of this technology relative to conventional activated sludge systems is attributed to the elimination of expensive separate secondary settlers and recirculation pumping systems, and reduction of land requirements by up to 80%.

4.9. Overall System Performance

Chemically enhanced pseudoliquified activated sludge bioreactor (CEPASB) technology is inclusive of the following processes and facilities:

- Bar screening of wastewater.
- Grit removal.
- Equalization tank—for wastewater equalization and reduction of facilities volume.
- Primary settling, for total suspended solids concentration in excess of 350.0 mg/L.
- Pumps and flow meter for pumping of wastewater from equalization tank into the pseudoliquified activated sludge bioreactor.
- Pseudoliquified activated sludge bioreactor—aeration tank-clarifier.
- Bioreactor with media and immobilized microorganisms.
- Tertiary polishing filter.
- Waste sludge measuring-tank or feed-pumps for waste activated sludge pumping.
- Unit for chemical reagent preparation and dosing with mixing chemical feed pumps for addition of ferric chloride solution, $FeCl_3$, or aluminum sulphate, $Al_2(SO_4)_3 \cdot 16\ H_2O$.
- Operation personnel and blowers, pumps, and control and measuring equipment as well as building and testing laboratory.
- Thickening of waste activated sludge.
- Sludge digestion facilities.
- Sludge dewatering facilities: sludge beds or frame Presses.
- Compost beds for sludge composting and use.
- Disinfection of the tertiary treated effluent (UV units or chlorination).

When chemically enhanced pseudoliquified activated sludge bioreactor (CEPASB) technology is applied for high-strength industrial wastewater treatment, multistage schemes are recommended.

4.10. Municipal and Industrial Wastewater Treatment—Process Applicability

The process schemes are based on a new combination of biological and chemically enhanced biological treatment technologies in the module reactors of the prefabricated system.

CEPASB technology is applied for municipal and industrial wastewater treatment. This technology is applicable for complete industrial and municipal wastewater treatment to meet surface discharge criteria for sensitive rivers as well as local industrial wastewater pretreatment for into municipal sewage system discharge.

Potential areas for utilization of the Chemically Enhance Pseudoliquified Activated Sludge Bioreactor (CEPASB) technology are:

- Municipal wastewater treatment plant for cities, towns, villages, resorts, and camping grounds.
- Industrial wastewater treatment plants for the food processing industries (i.e., meat, poultry, dairy, fish, and vegetable).
- Existing municipal and industrial wastewater treatment plants upgrade for process parameter optimization, performance, and operation enhancement, effluent quality improvement, as well as nitrogen and phosphorus removal, and reduction of waste activated sludge.

5. COMPUTER MODELING

5.1. Model Descriptions

The computer model used to simulate the pseudoliquified activated sludge system is the Activated Sludge Model No. 2 (ASIM), developed by the International Water Quality Association (IAWQ). The IAWQ model has been the basis of several commercial packages (i.e., GPS-X (Hydromantis Inc., ON) and Biowin^{R} (EnviroSim, ON) which are more user-friendly than ASIM. The model allows for dynamic simulation of combined biological processes for organics as well as nitrogen and phosphorous removal in suspended growth bioreactors (i.e., activated sludge (AS)).

ASIM accounts for particulate as well as soluble species or organic matter, nitrogen, and phosphorous. It also characterizes organic matter on the basis of biodegradation kinetics (i.e., readily biodegradable soluble and particulate organics) and relatively inert soluble and particulate organics. It also allows for inclusion of mineral particulate solids in the influent to treatment plants, as well as generation of such solids in the context of phosphorous precipitation. The most salient features of the model, in addition to dynamic simulations of extended operational periods are:

- Allowance for temperature-dependent variation of kinetic coefficients.
- Incorporation of both chemical and biological phosphorous removal processes.
- Addition of process controls to activated sludge systems.

It must be emphasized that the model requires extensive wastewater characterization data, sufficient to close mass balances on soluble, and particulate total organics, nitrogen,

and phosphorous. Wastewater characterization may involve expensive techniques such as respirometry. The model is devised for completely mixed suspended growth bioreactors treating non-inhibitory wastes in systems operating within the typical conditions for municipal wastes and, therefore, is not capable of predicting performance of secondary clarifiers, particularly with respect to suspended solids removal.

The model may be adapted to simulate the pseudoliquified activated sludge system, as the system employs a combination of completely mixed zones, as well as sludge zones with sufficient biological activity and biomass densities approaching fixed-film systems.

5.2. Wastewater Characterization

The model used in this study characterizes organic matter based on COD as follows:

- S_F: the soluble readily biodegradable.
- S_{AC}: soluble fermentable biodegradable (i.e., volatile fatty acids).
- S_I: inert soluble COD.
- X_I: inert particulate COD.
- X_S: biodegradable particulate.
- X_{BM}: heterotrophic biomass.
- $X_{BM\ AUTO}$: autotrophic biomass.

Statistical correlations between the soluble and total BOD and COD in the grit removal tank effluent, the aeration tank effluent, and the bioreactor effluent were developed and used for modeling. The observed BOD: COD ratio in the grit removal tank effluent was 0.6:1, which reflected good waste biodegradability. The typical BOD: COD ratio for municipal wastewater is 0.5:1.

Because the grit removal effluent represented the influent to the biological treatment system for most of the experimental program, the analysis of soluble fractions of this wastewater is particularly important. Based on the developed statistical correlations the following observations are noteworthy:

- Soluble fraction of BOD and COD in the grit removal tank effluent was approximately 35% to 40%.
- Grit removal BOD was mostly in the 80 to 180 mg/L range whereas COD varied from 150 to 300 mg/L.
- Soluble biodegradable organic matter in the system effluent was mostly in the range of 2.0 to 6.0 mg/L.
- Soluble COD in the system effluent was mostly in the range of 25 to 60 mg/L and, therefore, the inert soluble COD was calculated as the difference between the soluble COD and the COD equivalent of the 2 to 6 mg/L BOD, based on the ratio established above, to be 21 to 50 mg/L, or approximately 14% to 17% of the total influent COD, and accordingly a 15% was assumed.
- Fermentable substrate although not directly measured, was estimated to be 5% of the total COD or approximately 12.5% to 15% of the influent soluble COD.
- Particulate COD was 60% to 65% of the total influent COD.
- Particulate COD in the aeration tank effluent was about 8% of the COD or 2 to 5 mg/L (i.e., non-biodegradable particulate COD is 1.5% of the total influent COD).
- Influent heterotrophic biomass was estimated at approximately 5% of total influent COD (ASIM Manual).

- Biodegradable particulate COD leaving in the system effluent was estimated at 21% of the total BOD or 0.4 to 1.2 mg/L BOD corresponding to 0.7 to 2.0 mg/L COD.

Average total nitrogen in the grit removal effluent was calculated by using a typical ratio of ammonia to total nitrogen of 0.65 to 0.70, and was found to be in the range of 27.4 to 48.1 mg/L with most of the data in the 30.5 to 42 mg/L range. Based on the data discussed above, the following two scenarios for wastewater characteristics represented the average and maximum strength of the wastewater (see Table 9.6).

5.3. Determination of Model Stoichiometric Coefficients

The stoichiometric coefficients for ASIM are based on material balances for soluble and particulate organics, nitrogen, and phosphorous. The two alternative wastewater characteristics were tested to check the applicability of a single set of coefficients to both alternatives. It is interesting to note that a unique set of stoichiometric coefficients adequately described both scenarios and therefore is deemed applicable for all simulations.

5.4. Process Modeling

5.4.1. System Modeling

The biggest challenge in modeling this pseudoliquified activated sludge system using the Activated Sludge Model No. 2 (ASIM), which is developed for suspended growth systems, was the conversion of the hybrid bioreactor clarifier zones with varying oxidation-reduction potentials conditions to equivalent bioreactor and clarification zones.

Based on the following system specifics presented in Section 4 of this chapter:

- Aeration column volume is 1.0 m^3.
- Clarifier column volume is 11.8 m^3 comprised of the following zones.
- Treated wastewater and effluent collection zones with a combined volume of 1.9 m^3.
- Total surface area of pseudoliquified activated sludge layers of 5.7 m^2 and an observed thickness of 0.7 m almost independent of loading conditions.
- Total active bioreactor system volume of 10.9 m^3 (11.8 m^3 clarifier column +1.0 m^3 aeration tank −1.9 m^3 effluent collection zone).

The following assumptions were made to model the integrated aeration tank-clarifier system:

- The aeration tank volume of 1.0 m^3 was fully aerobic and its entire volume was available for biodegradation as a completely mixed system.
- Clarification area was 5.7 m^2 with a total volume of 3.95 m^3.
- Owing to mixing limitations in the activated sludge layers in the clarifier column and ensuing mass transfer limitations, the equivalent mixing efficiency was 85% (i.e., the equivalent bioreactor volume in the column) calculated as 0.85*(11.8-3.95-1.9), was 6.0 m^3.
- Owing to the high recirculation ratio of return sludge to influent flow of 800%, 60% of the active bioreactor volume in the clarifier column was aerobic and the rest was anoxic (i.e., 3.5 m^3 aerobic and 2.5 m^3 anoxic).
- Total active bioreactor volume of 7.0 m^3, comprised of the 6.0 m^3 in the clarifier columns and the 1.0 m^3 aeration tank.
- Biomass in the system was mostly in the pseudoliquified layers and the aeration column with minimal solids in the clarified zones and effluent collection chambers, thus implying that the total

Table 9.6
Raw wastewater characteristics used for modeling

Conversion Factors										
Stoichiometric Constants										
	Nitrogen		Phosphorus		TSS			fSI =	0	
SI = inert soluble COD	iNSI =	0.05	iPSI =	0.01	iTSSXI =	0.75		YH =	0.63	
SF = soluble subtrate	iNSF =	0.05	iPSF =	0.01	iTSSXS =	0.75		fXI =	0.1	
XI = inert particulate COD	iNXI =	0.07	iPXI =	0.01	iTSSBM =	0.9		YPAO =	0.63	
XS = particulate substrate	iNXS =	0.07	iPXS =	0.01				YrPO4 =	0.4	
BM = biomass	iNBM =	0.08	iPBM =	0.016				YrPHA =	0.2	
fSI = fraction of inert COD in XS								YAUT =	0.24	
Y = yield coefficients										
Yr = storage requirements										
fXI = fraction of inert COD generated in biomass lysis										
Mass Balance Check										
			TKN	Phosphate						
Soluble			35.25	4.35		SI	30		SPO4	3.3
Particulate			12.75	1.89		SF	75		SNH4	30
Estimated total			48	6.24		XI	30			
Average actual total			44	7		XS	135			
Error			−9.09%	10.86%		BM	15			

Stoichiometric Matrix for Dissolved Components										
	SO2	SF	SA	SNH4	SNO3	SPO4	SI	SALK	SN2	SFe
Process										
1 Aerobic hydrolysis		1		0.02		0		0.0014286		
2 Anoxic hydrolysis		1		0.02		0		0.0014286		
3 Anaerobic hydrolysis		1		0.02		0		0.0014286		

Heterotrophic Organisms: XH									
4 Growth on SF	−0.587302	−1.58730		−0.000634		−0.000127		−3.92E-05	
5 Growth on SA	−0.587302		−1.587302	−0.08		−0.016		0.0199	
6 Denitrification with SF		−1.58730		−0.000634	−0.20535	−0.000127		0.0146	0.2054
7 Denitrification with SA			−1.587302	−0.08	−0.20535	−0.016		0.0345	0.2054
8 Fermentation		−1	1	0.05		0.01		−0.012537	
9 Lysis				0.01		0.006		0.0004	
Phosphorus Accumulation Organisms (PAO): XPAO									
10 Storage of XPHA			−1			0.4		−0.00373	
11 Storage of XPP	−0.2					−1		0.0484	
12 Aerobic growth	−0.587302			−0.08		−0.016		−0.00494	
13 Lysis of XPAO				0.01		0.006		0.0004	
14 Lysis of XPP						1		−0.048387	
15 Lysis of XPHA			1					−0.015625	
Nitrifying Organisms (Autotrophic Organisms): XAUT									
16 Growth		−18.04167			−4.246667	4.1667	−0.016		−0.600178
17 Lysis					0.01		0.006		0.0004
Simultaneous Precipitation of Phosphorus with Ferric Hydroxide ($Fe(OH)_3$):									
18 Precipitation							−1		0.0484
19 Redissolution							1		−0.048387
Oxidation of Iron									
20 Iron oxidation		−0.14						0.018	−1

Table 9.6 (*Continued*)

Conversion Factors										
Stoichiometric Constants										
	Nitrogen		Phosphorus		TSS			fSI =	0	
Stoichiometric Matrix for Particulate Components										
	XI	XS	XH	XPAO	XPP	XPHA	XAUT	XTSS	XMeOH	XMeP
Process										
1 Aerobic hydrolysis		−1						−0.75		
2 Anoxic hydrolysis		−1						−0.75		
3 Anaerobic hydrolysis		−1						−0.75		
Heterotrophic Organisms: XH										
4 Growth on SF			1					0.9		
5 Growth on SA			1					0.9		
6 Denitrification with SF			1					0.9		
7 Denitrification with SA			1					0.9		
8 Fermentation										
9 Lysis	0.1	0.9	−1					−0.15		
Phosphorus Accumulation Organisms (PAO): XPAO										
10 Storage of XPHA					−0.4	1		−0.692		
11 Storage of XPP					1	−0.2		3.11		
12 Aerobic growth				1		−1.587302		−0.052381		
13 Lysis of XPAO	0.1	0.9		−1				−0.15		
14 Lysis of XPP					−1			−3.23		
15 Lysis of XPHA						−1		−0.6		
Nitrifying Organisms (Autotrophic Organisms): XAUT										
16 Growth							1	0.9		
17 Lysis	0.1	0.9					−1	−0.15		
Simultaneous Precipitation of Phosphorus with Ferric Hydroxide (Fe(OH)3):										
18 Precipitation								1.42	−3.45	4.87
19 Redissolution								−1.42	3.45	−4.87
Oxidation of Iron										
20 Iron oxidation								1.91	1.91	

volume for biosolids accumulation is 5.0 m^3, comprised of the 1.0 m^3 in the aeration tank plus the 4.0 m^3 in the sludge layers.
- Biological solids residence time (SRT) is thus 40% of the total SRT in the system (5.0 m^3/12.8 m^3).
- Chemicals for phosphorous removal were added to the aeration tank.

With the above assumptions, the integrated aeration tank-clarifier was modeled as a sequence of the following two independent systems:

1. ***Aerobic System***
   Reactor volume = 4.5 m^3
   Dissolved oxygen concentration = 3.0 mg/L
   Clarifier volume = 2.5 m^3
   Ratio of return activated sludge flow to influent flow = 8:1
   Aerobic SRT = 64% of biological SRT
   Influent: Actual system influent (i.e., grit removal tank effluent)
2. ***Anoxic System***
   Reactor volume = 2.5 m^3
   Dissolved oxygen concentration <0.5 mg/L; mass transfer coefficient = 3.0/day
   Clarifier volume = 2.5 m^3
   Ratio of return activated sludge flow to influent flow = 1:1
   Aerobic SRT = 36% of biological SRT
   Influent: settled aerobic effluent

It was assumed that all particulate species except for 20 mg/L of volatile suspended solids are removed in the clarifier of the aerobic system (i.e., only soluble COD, N, and P fractions are fed to the second system).

Because all organic matter would be consumed in the aerobic system, the organic matter required for denitrification would be supplied by the decay of biomass, and would be treated as a concentrated second influent to the system based on a flow rate of 1.0 m^3/day. Such a flow rate would represent less than 3% of the system influent flow (i.e., does not significantly change the hydraulic retention time in the system) and, therefore, would not impact predictions. A first-order biomass decay coefficient of 0.15/day and a conversion factor of 1.1 g COD/g VSS decaying was used to calculate organic load to the anoxic system.

5.4.2. Scenarios Modeled

During the 284-days commissioning and operational testing of the pilot-plant system, four periods were operated at steady-state conditions. In a conventional suspended growth system, 2 to 3 turnovers of the mean SRT are required to achieve steady-state conditions. In this system, however, because of the high mixed liquor suspended solids concentration and the fact that biological SRT is only 40% of the total SRT, steady-state conditions were mostly indicated by the stability in effluent concentrations. The following operational conditions were modeled:

1.0 Flow = 40.8 m^3/day, HRT = 6.4 h, SRT = 37.6 days, T = 13°C to 15°C
2.0 Flow = 40.8 m^3/day, HRT = 6.4 h, SRT = 57.4 days, T = 13°C to 20°C

3.0 Flow = 40.8 m^3/day, HRT = 6.4 h, SRT = 37.6 days, T = 16°C to 24°C
4.0 Flow = 54.7 m^3/day, HRT = 4.8 h, SRT = 17 days, T = 20°C to 30°C

Because chemical addition for P removal was initiated during Stage 4, this scenario was modeled both using and excluding iron addition at an iron dose of 15 mg/L. Details of the influent and effluent characteristics for each of the operational stages listed above are obtained during the mobile pilot treatment plant operation period.

5.4.3. Modeling Results

A comparison of the effluent quality parameters with the model predictions is presented in Table 9.7. It should be emphasized that the biokinetic coefficients used to model the system are well within the typical values for municipal wastewater treatment, except for the decay coefficient of 0.15/day, which is at the upper limit of accepted values. The rationale for using a relatively high biomass decay coefficient is that biomass decay rates increase with SRT and the very long SRTs in this system (approaching SRTs for aerobic sludge digestion) is thus reflected by a high decay coefficient. Additionally, the high decay coefficient is also justified by the observed sludge yields, which are about 30% to 50% below typical values for municipal wastes.

Predicted mixed liquor suspended solids concentrations reported in Table 9.7 are the weighted averages of the concentrations in the aerobic and anoxic bioreactors. Furthermore, although the model is COD-based and does not provide predictions of BOD, soluble BOD may be estimated from the sum of the readily biodegradable substrate COD and the fermentable COD. It is apparent that the model predicted soluble effluent species (i.e., BOD, ammonia-nitrogen, nitrates-nitrogen, orthophosphates-phosphorous) as well as MLSS and MLVSS concentrations more accurately than soluble COD. The primary reason for this discrepancy in behavior is that the nonbiodegradable or inert soluble COD, was most likely the most variable wastewater component, as the Kiev wastewater treatment plant treats a mixture of industrial and municipal wastewaters.

The predictions for the highest SRT scenario of 57.4 day are not included in Table 9.7, as they did not yield satisfactory predictions of biomass, which in turn would have resulted in much higher denitrification rates than observed experimentally. This period was characterized by high influent SS concentrations, which may be owing to experimental analytical errors.

The sensitivity of the model predictions to wastewater temperature was tested by varying the temperature for Stage 1 from an average temperature of 13.7°C to 20°C. Predicted effluent ammonia concentrations decreased drastically from 8 to 2.5 mg/L, whereas nitrate concentrations rose from 0.26 to 1.6 mg/L and MLSS concentration increased by 5% to 6587 mg/L. Temperature effects can also be discerned by comparing the system performance during Stages 1 and 3, which were identical with respect to hydraulic and organic loadings and SRT, and varied only in temperature. It appears from the results of the modeling that temperatures of 15°C and above are conducive to good nitrification. The sensitivity of the reactor performance with respect to nitrification to temperature changes is more pronounced than conventional activated sludge systems and biological nutrient removal processes, owing to the very short hydraulic retention time in the system.

Table 9.7
Comparison of experimental data and model predictions for various operational conditions

Stage	Soluble COD		Soluble BOD		Ammonia-nitrogen		Nitrates-nitrogen		Orthophosphates-P		MLSS		Volatile Fraction	
No.	Measured	Predicted	Measured	Predicated	Measured	Predicated	Measured	Predicated	Measured	Predicated	Measured	Predicated	Measured	Predicated
1	N/A	77	N/A	63	8.0	10.5	1.1	0.3	4.6	5.7	6445	6285	0.78	0.72
3	51.0	28.6	3.7	2.30	1.9	1.95	7.9	3.6	4.4	4.1	6095	6302	0.75	0.82
4A	47.2	21.9	3.4	3.03	2.1	1.00	2.4	5.2	2.8	3.0	6506	5945	0.77	0.78
4B	47.2	21.9	3.4	3.03	2.1	1.00	2.4	5.2	0.46	0.34	6506	6000	0.77	0.77

N/A denotes not available.
4A refers to operation without iron addition.
4B refers to operation with 15 mg/L as Fe added.

6. SUMMARY AND RECOMMENDATIONS

These state-of-the-art technologies and treatment plants in Ukraine, Russia, and Belarus generally meet the world standards for treatment and are protected by the patents (10–16). Full-scale facilities are serving communities with population ranging from 3,500 to 1,500,000 in the three countries (22, 29, 30). The readers are referred to literature (32–38) for more information regarding the process and its recent developments.

The installed systems have been in successful operation from 5 to 15 years, treating a broad range of wastewater flows from 100 to 400, 000 m^3/day.

This technology has several advantages over conventional activated sludge systems:

1. Compactness as a result of reduction of land requirements by up to 80% to 90%.
2. Increase in oxygen utilization efficiency owing to increased submergence depth.
3. Elimination of expensive separate settlers and recirculation pumping stations; and reduction of both capital and operating costs, owing to lower energy consumption and sludge production.

Because previous operating and research experiences with the pseudoliquified activated sludge bioreactor have not focused on phosphorous removal to <1 mg/L, recent development of the technology addressed achievability of higher P removal efficiencies as well as a better understanding of the fundamental processes occurring within the system. Thus, a mobile pilot unit utilizing chemical solution for P removal was tested in Ukraine. This chemically enhanced pseudoliquified activated sludge bioreactor (CEPASB) was built and tested, over a wide range of operational conditions to establish optimum design parameters and optimum operating conditions. The results of this pilot study are summarized herein.

From a nutrient standpoint, the optimum hydraulic loading rate was observed to be 2.28 m^3/h (54.72 m^3/day), (from August 08, 2001 to August 30, 2001), with the following ranges of nitrogen concentrations: plant influent: total nitrogen, N—27.36 to 48.12 mg/L; ammonia-nitrogen, $NH_4^{+-}N$—13.3 to 34.5 mg/L; nitrites, NO_2-N—0.00 to 0.18 mg/L; nitrates, NO_3-N—0.00 mg/L; aeration tank effluent: total nitrogen—2.61 to 3.91 mg/L; ammonia-nitrogen—0.27 to 2.1 mg/L; nitrites-N—0.00 to 0.06 mg/L; and nitrates-N—0.00 to 2.9 mg/L.

The following pilot plant process parameters as achieved at the hydraulic loading of 2.28 m^3/h (54.72 m^3/day), are recommended for technology design and application:

1. Hydraulic loading—2.28 m^3/h (54.72 m^3/day).
2. Wastewater retention time (wastewater contact with activated sludge)—4.78 hours.
3. Solid residence time, SRT—12 to 17 days.
4. Optimum organic loading rates based on total and soluble BOD and COD of 0.14 mg BOD_5/mg VSS/day, 0.05 mg $SBOD_5$/mg VSS/day, 0.235 mg COD/mg VSS/day, and 0.054 mg SCOD/mg VSS/day respectively.
5. Sludge yields varied from 0.07 to 0.68 g TMSS/COD (0.257 g TMSS/g COD) and from 0.03 to 0.65 g TMSS/g BOD_5, with an average yield of 0.176 g TMSS/g BOD_5.
6. At the optimum dose of ferric chloride determined to be in the range of 72.5 or 15.0 mg/L (Fe) to 80.9 or 16.7 mg/L (Fe), the standard phosphorus criterion for surface discharge (i.e., <1 mg/L) were achieved as follows: Orthophosphates, PO_4-P—0.29 to 0.5 mg/L, total phosphorus, P—0.56 to 0.76 mg/L.

It is anticipated that the optimum dose of ferric chloride at full-scale wastewater treatment plant will be much less than observed in this study, owing to accumulation of ferric chloride particles into suspended layers of activated sludge.

During the mobile pilot plant testing period, solids concentration in the aeration tank were used in conjunction with the 30 minutes settling data to estimate sludge volume index (SVI) and assess settleability. The values of sludge volume index (SVI) during testing (without ferric chloride, $FeCl_3$, addition) were in the range 50 to 123 mL/g. With addition of ferric chloride, settleability of activated sludge improved as reflected by a drop in sludge volume index (SVI) from 70 to 100 mL/g to 40 to 50 mL/g.

A computer model of this technology using Activated Sludge Model No. 2 (ASIM) was developed, calibrated, and tested against plant data. The system was modeled as a combination of two independent activated sludge systems, one aerobic and the other anoxic. The biological solids residence time (SRT) was modeled as being 40% of the system SRT mainly attributable to accumulation of solids in specific zones. The model predicted the experimentally measured soluble BOD, ammonia-nitrogen, nitrate-nitrogen, nitrite nitrogen, orthophosphates as phosphorous, mixed liquor total and volatile suspended solids accurately with discrepancies well within accuracy of measurement. However, soluble COD was predicted reasonably well but deviations from actual measurements were observed most probably attributable to variations in the influent concentrations of soluble inert COD. The model confirmed the sensitivity to temperature observed experimentally, with substantial loss of nitrification at temperatures below 13°C.

NOMENCLATURE

a_i, a_{act} = activated sludge concentrations, g/L
T = solid residence time, day
T = hydraulic retention time, h
P_i = yield of activated sludge, mg/L
ρ_{org} = specific rates of oxidation of organic nitrogen, mgN/($g_{MVSS} \cdot h$)
ρ_{nitr}, ρ_{den} = specific rates of nitrification, denitrification, mgN/($g_{MVSS} \cdot h$)
ρ_{tot} = specific rates of total nitrogen removal, mgN/($g_{MVSS} \cdot h$)
s = Nonvolatile or inert fraction of suspended solids
I = sludge volume index, mL/g
N_{tot}, N_{org} = total nitrogen, organic nitrogen concentrations, mgN/L
N_{NH4}, N_{NO2}, N_{NO3} = ammonia-nitrogen, nitrites, nitrates concentration, mgN/L

REFERENCES

1. USEPA, EPA releases "flexible" Great Lakes guidelines. *Journal of Water Environment and Technology, Water Environment Federation*, 25 (1995).
2. S. V. Jakovlev, The problems of wastewater treatment, *Water Supply and House Technics J., Construction Literatures Edition*. 7, 2–3 (1994).
3. J. Barica, Explosion of environmental issues? *Journal of Water Quality Research Canada*, 30, 2, 151–153 (1995).
4. J. B. Copp and K. L. Murphy, Estimation of the active nitrifying biomass in activated sludge, *Journal of Water Research*, 29. 8, 1855–1862 (1995).

5. W. Gujer, M. Henze, T. Mino, T. Matsuo, M. C. Wentzel, and G. V. R. Marais, The activated sludge model # 2: biological phosphorus removal, *Water Science Technology*, 31. 2, 1–11 (1995).
6. M. V. Wable and C. W. Randal, Investigation of hypothesized anaerobic stabilization mechanisms in biological nutrient removal systems, *Water Environment Research*, 29. 8, 1855–1862 (1994).
7. G. Tchobanoglous and F. L. Burton, Advanced wastewater treatment, In: *Wastewater Engineering: Treatment, Disposal and Reuse*, 3rd ed., pp. 701–721, Metcalf & Eddy, McGraw-Hill, (1991).
8. A. Sverdlikov, M. Zemljak, E. Volodarsky, D. Haycock, and A. Lugowski, Ukraine develops modular plants, *Journal of Environmental Science and Engineering*, 68 (1994).
9. M. M. Zemljak, A. I. Sverdlikov, A. A. Bondarev, and A. A. Sverdlikov, Wastewater treatment plants of column-type. *Journal of Water Supply and House Technics, Construction Literatures Edition*. 9, 4–6 (1994).
10. A. I. Sverdlikov, M. M. Zemljak, and M. E. Kigel, *Aeration Tank of Column-Type*. USSR Patent 604287, C02C 1/2, F 3/00, August, Vol. 29 (1981).
11. M. M. Zemljak, A. I. Sverdlikov, A. A. Bondarev, A. A. Sverdlikov, A. V. Ershov, V. N. Shvetsov, and I. N. Mjasnikov, *Installation for the Treatment of Wastewater and Natural Water*. USSR Patent 1721028, C02 F 3/00, March, Vol. 11 (1992).
12. A. I. Sverdlikov, M. M. Zemljak, and M. E. Kigel, *Combine Installation for Treatment of Industrial and Wastewater*. French Patent 2451892, C02F 1/02, F 3/009, April, Vol. 4/60, pp. 14 (1981).
13. A. I. Sverdlikov, M. Zemljak, E. Volodarsky, A. A. Sverdlikov, and D. Haycock, Deeper by design: biological treatment of wastewater in column bioreactors with Pseudoliquified layers of activated sludge and fixed film microorganisms. *Journal of Water Quality International*, November/ December, 47–48 (1997).
14. M. M. Zemljak, A. I. Sverdlikov, and A. A. Sverdlikov, *Aeration Tank-Clarifier of Column-type*. USSR Patent 1530573, C02 F 3/00, December, Vol. 47 (1989).
15. M. M. Zemljak, A. I. Sverdlikov, and A. A. Sverdlikov, Investigation of the factors that influence the effectiveness of the aeration processes. *Science and Technique in the Municipal Economy*, Kiev, Ukraine, 74 (1990).
16. M. M. Zemljak, A. I. Sverdlikov, A. A. Bondarev, A. A. Sverdlikov, A. V. Ershov, V. N. Shvetsov, and I. N. Mjasnikov, *Aeration Tank-Clarifier of Column-type*, USSR Patent 1638122, C02 F 3/00, March, Vol. 12 (1991).
17. M. M. Zemljak, A. I. Sverdlikov, A. A. Bondarev, and A. A. Sverdlikov, Wastewater treatment plants of column-type. *Journal of Water Supply and House Technics, Construction Literatures Edition*, 9, 4–6 (1994).
18. M. M. Zemljak, A. I. Sverdlikov, and A. A. Sverdlikov, Technology of complete biological wastewater treatment. *Science and Technique in the Municipal Economy*, Kiev, Ukraine, 3 (1994).
19. S. V. Jakovlev, A. A. Sverdlikov, G. P. Scherbina, M. M. Zemljak, and A. I. Sverdlikov, Technology of complete biological wastewater treatment the modular combined column and gallery bioreactors. *Ecology, Technology, Economy of Water Supply and Sewerage International Congress*, Yalta, Ukraine, April 15–17, pp. 33–34 (1997).
20. S. V. Jakovlev, A. A. Sverdlikov, A. I. Sverdlikov, and M. M. Zemljak, Technology of complete biological treatment of wastewater for nutrient removal in modular combined column and gallery bioreactors. *Water: Ecology and Technology, Second International Congress, ECWATECH-96*, Moscow, September 17–21, pp. 295 (1996).
21. A. A. Sverdlikov, M. M. Zemljak, E. Volodarsky, A. I. Sverdlikov, and D. Haycock, Deeper by Design: Biological treatment of wastewater in column and gallery bioreactors with Pseudoliquified layers of activated sludge and fixed film microorganisms. *Water Quality International*, November/ December, pp. 47–48 (1997).

22. S. V. Jakovlev, A. A. Sverdlikov, G. P. Scherbina, A. I. Sverdlikov, M. M. Zemljak, and D. Haycock, Cost effective technology of complete biological wastewater treatment in the column and gallery bioreactors with pseudoliquified layers of activated sludge and fixed film microorganisms. *Water: Ecology and Technology, Third International Congress, ECWATECH-98*, Moscow, May 26–30, pp. 331–332 (1998).
23. A. A. Sverdlikov, Complete biological wastewater treatment for nitrogen removal. *Synopsis of a Thesis, NIIVODGEO*, Moscow, pp. 27 (1996).
24. M. G. Djikija, Municipal wastewater treatment by one stage of nitrification and denitrification. *Synopsis of a Thesis, NIIVODGEO*, Moscow, pp. 26 (1986).
25. N. M. Ilinskaja, Nitrification of nitrogen contaminated wastewater. *Synopsis of a Thesis, NIIVODGEO*, Moscow, pp. 25 (1990).
26. M. M. Zemljak, Development and research of aeration tank-clarifier for municipal wastewater treatment. *Synopsis of a Thesis, KISI*, Kiev, pp. 27 (1975).
27. N.V. Muravjova, Two-stage wastewater treatment in the aeration tanks. *Synopsis of a Thesis, NIIVODGEO*, Moscow, pp. 27 (1996).
28. A.I. Sverdlikov, Aeration tanks with the phase's division of municipal wastewater treatment. *Synopsis of a Thesis, KISI*, Kiev, pp. 27 (1986).
29. S. V. Jakovlev, A. A. Sverdlikov, G. P. Scherbina, A. I. Sverdlikov, M. M. Zemljak, and D. Haycock, New technology of complete biological wastewater treatment in the column and gallery bioreactors with pseudoliquified layers of activated sludge and fixed film microorganisms. *Water: Ecology and Technology, Fourth International Congress, ECWATECH-2000*, Moscow, May 26–30 (2000).
30. A. I. Sverdlikov, G. P. Scherbina, M. M. Zemljak, and A. A. Sverdlikov, Development and implementation of a mobile pilot plant for industrial and municipal wastewater treatment. *Ecology, Technology, Economy of Water Supply and Sewerage International Congress*, Yalta, Ukraine, May 15–17, pp. 35–37 (2001).
31. K. H. Rosenwinkel, F. Obenaus, and M. Wichern, Performance of biological nitrogen and phosphorous removal in four large WWTPs in Germany. *Water Science and Technology*, 41, 9, 233–240 (2000).
32. R. Abdul-Rahman, H. Ozaki, and N. Zainola, Expanded bed column bioreactor performance and hydrodynamics, *The Canadian Journal of Chemical Engineering*, 867–874 (2004).
33. T. P. West and B. Strohfus, Polysaccharide production by immobilized Aureobasidium pullulans cells in batch bioreactors, *Microbiological Research*, 156, 3, 285–288 (2001).
34. N. Wery, U. Gerike, A. Sharman, B. Chaudhuri, D. W. Hough, and M. J. Danson, Use of a packed column bioreactor for isolation of diverse protease producing bacteria from Antarctic soil, *Applied Environmental Microbiology*, 69, 3, 1457–1464 (2003).
35. A. F. Hsu, E. Wu, S. Shen, T. A. foglia, and K Jones, Immobilized lipoxygenase in a packed bed column bioreactor: continuous oxygenation of linoleic acid, *Biotechnology and Applied Biochemistry*, 30, 245–250 (1999).
36. L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas, (eds.), *Biological Treatment Processes*, Humana Press, Totowa, NJ, 818 pp. (2009).
37. L. K. Wang, N. K. Shammas, and Y. T. Hung, (eds.), *Advanced Biological Treatment Processes*, Humana Press, Totowa, NJ (2009).
38. L. K. Wang, N. K. Shammas, and Y. T. Hung, (eds.),. *Biosolids Treatment Processes*, Humana Press, Totowa, NJ, 820 pp. (2007).

10
Upflow Sludge Blanket Filtration

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CONTENTS

Abstract The use of upflow sludge blanket filtration (USBF) covers the whole spectrum of chemical and biological treatment of water using the agglomeration processes for transformation of colloidal and dissolved impurities in water into separable floc suspension. The technical applications of USBF involve a wide range of treatment of underground and surface water for communal and industrial use, and municipal, industrial, and agricultural biological wastewater treatment.

The fluidized bed filtration is a complex phenomenon based on a number of principles. In addition to the exposition of those principles, this chapter discusses the principles of integrated USBF reactor's design, gives examples of USB technology implementation, explains the advanced wastewater treatment systems and works out a design example of advanced treatment systems, the upgrading of a classical municipal WWTP.

Key Words Reactor design • upflow sludge blanket filtration • USBF technology • wastewater treatment.

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_10 © Humana Press, New York, NY 2009

1. INTRODUCTION

One of the most innovative physical unit operations in water treatment is the separation of floc suspension by fluidized bed filtration, known in water treatment by the name upflow sludge blanket filtration (USBF). The innovation radius of USBF covers the whole spectrum of chemical and biological treatment of water using the agglomeration processes for transformation of colloidal and dissolved impurities in water into separable floc suspension. The technical applications of USBF involve a wide range of chemical treatment of underground and surface water for communal and industrial use, and municipal, industrial, and agricultural biological wastewater treatment.

The industrial applications of USBF technology began by implementation of the fluidized bed filtration for chemical treatment of surface water for potable use. The principles of USBF in chemical treatment of water and various design of separators based on this technology have been disclosed and developed in the early 1950s (1, 2).

In the early 1970s the next modification of USBF technology was developed. It was based on the principle of fluidized bed filtration in the form of partially fluidized bed (3). This innovation extends the application of the USBF technology to the biological treatment of wastewater and by the end of the eighties it had further developed to its present form (4) and applied in a number of wastewater treatment plants.

The innovative potential of the new principle in water treatment technology culminated in the development of a new principle of fluidized bed filtration at the beginning of the new millennium in the form of COMBI USBF (5), which introduced a new dimension into the biological wastewater treatment. The innovation radius of the COMBI USBF technology is not limited to design of new treatment facilities, but it is influencing even the basic concept of the water management in general.

For example, the high effluent water quality obtainable by USBF technology plays a decisive role in wastewater reclamation philosophy, providing new concepts for solving the deficit in water resources. The small dimensions and compactness of the integrated USBF reactors and their easy operation and control together with high reliability provides new dimensions in design of small technological units as a basis for decentralized sewage systems with small local wastewater reclamation facilities. The capacity of USBF for upgrading of existing wastewater treatment plants represents the third major innovation radius.

2. THEORETICAL PRINCIPLES OF FLUIDIZED BED FILTRATION

The fluidized bed filtration is a complex phenomenon based on a number of principles. To make the presentation of it more lucid, we have to start with individual exposition of those principles, however, only in extent necessary for our purpose.

2.1. Hydrodynamic Similarity and Dimensionless Numbers

The fluid motion can be described by Navier–Stokes Equations, and the solution of those equations with proper boundary conditions would give required results for the case under study. Alas, Navier–Stokes equations are nonlinear partial differential equations, so that their general solution cannot be obtained analytically, and they are four-dimensional

(three-dimensional space + time), so that even the numerical solution of cases of practical importance with sufficient precision is practically out of the ability of existing computers (there is moreover the complication owing to the instability of the numerical solution). The most practical way is therefore based on the transfer of experimental results from one case to another. This transfer is based on the theorem of similarity—"two structures behave similarly if they are geometrically, kinematically, and dynamically similar" (6). For the similarity purpose, the dimensionless entities, also called groups, numbers, or criteria, are used. In the dimensionless numbers there are grouped together different geometrical, kinematical, and dynamical parameters characterizing the case under study in such a way, that the dimensions of used parameters are mutually canceled. The full list of dimensionless groups can be found, e.g., in Reference (6). For our purpose the following dimensionless numbers are important:

Reynolds number Re. It is the best-known dimensionless number in hydrodynamics. The nature of fluid flow depends on the *Re* value, for small *Re* values it is laminar, for great *Re* values it is turbulent. The general definition of *Re* is

$$Re = \rho v L/\mu \tag{1}$$

where ρ and μ are the specific mass and the kinematic viscosity of liquid (note that μ/ρ is the dynamic viscosity), v is the characteristic velocity of liquid and L is the characteristic length for given case.

Archimedes number *Ar*. Its general definition is

$$Ar = (\rho_s - \rho)\rho g L^3/\mu^2 \tag{2}$$

where ρ_s is the specific mass of solid and g is the gravitation constant.

Drag coefficient f_d. Its general definition is

$$f_d = (F/A)/(\rho v^2/2) \tag{3}$$

where F is the drag force and A is a characteristic projected area.

Euler number *Eu*. Its general definition is

$$Eu = (\text{grad } p)L/\rho v^2 \tag{4}$$

where p is a pressure due to the liquid flow.

2.2. *Characteristics of Granular Porous Medium*

The granular porous medium is a medium formed by discrete particles—a typical example is a layer of sand. It can be described by some parameters, which are used as a standard for this purpose (7). They are:

Particle diameter D. In case the particles forming porous medium are spheres of the same size, it is the diameter of this sphere. In case the particles are identical but they are not spheres, it is the diameter of equivalent sphere, i.e., the diameter of the sphere with the same volume as a particle. In case the particles are different, it is a mean diameter value in a given place.

Surface factor Φ. It correlates the properties of nonspherical particle to the properties of a sphere. For spherical particle $\Phi = 1$, for nonspherical particle $\Phi < 1$. Geometrical Φ

expresses the ratio of the surface of a sphere to the surface of a particle with the same D. Hydrodynamical Φ express the value given by substitution of $D\Phi$ for D in a correlated dimensionless criteria providing the same hydrodynamic results as for the sphere with same D. In fixed granular porous medium for laminar flow and not too small D, the geometrical Φ and hydrodynamical Φ are equal; however, in some cases they can be different. This difference can have various reasons as e.g., the capillary forces, the preferred particle orientation in certain flow regimes, the ratio of boundary layer to surface roughness, etc.

Porosity ε. It is the ratio of pore volume to total volume of porous medium in a given place. It is also called voidage. The ratio of total volume of particles to the total volume of porous medium in given place is $1 - \varepsilon$. The range of ε values for fixed porous medium is limited from the value for most dense packing to the value for most loose packing of particles. For most loose arrangement of spheres (cubic packing) theoretically $\varepsilon = 0.476$, for nonspherical particles it can be even greater.

Specific surface P. It is the total surface of particles in volume unit of porous medium. For granular bed formed by identical particles the specific surface is

$$P = 6(1 - \varepsilon)/D\Phi \tag{5}$$

Distribution function $f_{\mathrm{D}}(D)$. It expresses the distribution of particles with respect to the particle diameter D. Fraction of particles having the diameter between D_1 and D_2 values ($D_1 < D_2$) corresponds to the integral of $f_{\mathrm{D}}(D)$ from D_1 to D_2. For medium from identical particles, the distribution function is a Dirac delta function at their D value.

2.3. Flow Through Fixed Porous Medium

For porous medium, the substitutions $v = V/\varepsilon$ and $L = \varepsilon/P$ in all dimensionless numbers are used, where V is apparent velocity of liquid, i.e., the flow divided by the total cross section (7). So

$$Re = \rho V/\mu P \tag{6}$$

$$Eu = (\mathrm{grad}\ p)\varepsilon^3/\rho V^2 P \tag{7}$$

Note that for linear flow along x-axis $(\mathrm{grad\ p}) = (\mathrm{d}p/\mathrm{d}x)$. For laminar and transient flow through fixed porous medium it was found that experimental results can be fitted by Equation

$$Eu = (5/Re)(1 + 0.073Re^{3/4}) \tag{8}$$

where the last term in the second bracket is due to the deviation from laminar flow. Using this equation, the mean value of P for fixed porous medium can be determined experimentally from the head loss measurement for a given V with knowledge of ε, ρ, and μ. Note that in a restricted range of *Re* it is for fixed porous medium often used with sufficient precision the power dependence

$$Eu = K_{\mathrm{R}} Re^n \tag{9}$$

where K_R and n are empirical constants depending on the chosen range. Generally, the value of power coefficient n lies between vales of 1 and 2 pertaining to laminar and fully developed turbulent flow, respectively.

According to above equations, the gradient of pressure increases with increasing flow velocity. Thus, the flow through porous medium has a tendency to cancel out the nonuniformity of flow, so that, e.g., the flow between point source and point sink is in porous medium broaden and in column in sufficient distance from both the flow is practically uniform.

2.4. Filtration

The term filtration is used in two meanings, the fist commonly used in hydraulics, the second commonly used in water treatment. The first is the flow of liquid through fixed porous medium, the second is the removal of suspended solids (SS) from the liquid by the flow of this liquid through porous medium. The removal of suspended solids (SS) can proceed by two different ways. In case the SS particles are greater then the medium pore dimension, it is simple straining. In case the SS particles are smaller then the medium pore dimension, they penetrate with flowing liquid into the porous medium and are caught inside it. It is this last type of filtration, which is important for fluidized bed filtration.

The gradual trapping of SS particles during the flow of liquid through porous medium results in gradual decrease of concentration of SS (8). The decrease of SS concentration C along the axis x of linear flow can be described by the equation:

$$C = C_0.e^{-ax}.f(ax) \tag{10}$$

where C_0 is the initial SS concentration,

$$a = (\mathrm{d}(C/C_0)/\mathrm{d}x)_{x=0} \tag{11}$$

is so-called filtration coefficient, and $f(ax)$ is the suspension nonhomogeneity function. For homogenous suspension $f(ax) = 1$, for nonhomogenous suspension $f(0) = 1$, $(\mathrm{d}f(ax)/\mathrm{d}(ax))_{x=0} = 0$, and with increasing value of ax the $f(ax)$ steadily increases $(\mathrm{d}^2 f(ax)/\mathrm{d}(ax)^2 > 0)$. For illustration, the theoretical curve for homogenous suspension and the experimental curve for nonhomogenous suspension with the same a value are shown in Fig. 10.1. The filtration coefficient depends on the nature of SS and parameters of porous medium and flow. The experimentally found dependence for fixed granular porous medium is

$$Ma = Re^{2.85}, \tag{12}$$

where Ma is the dimensionless filtration criterion

$$Ma = K_a P^2/a^2\varepsilon^2 \tag{13}$$

and K_a is a constant depending of SS nature. The greater value of filtration coefficient a, the thinner layer of porous medium is sufficient for required removal of suspension. From above results that filtration coefficient a increases with decreasing V and increasing P. It also

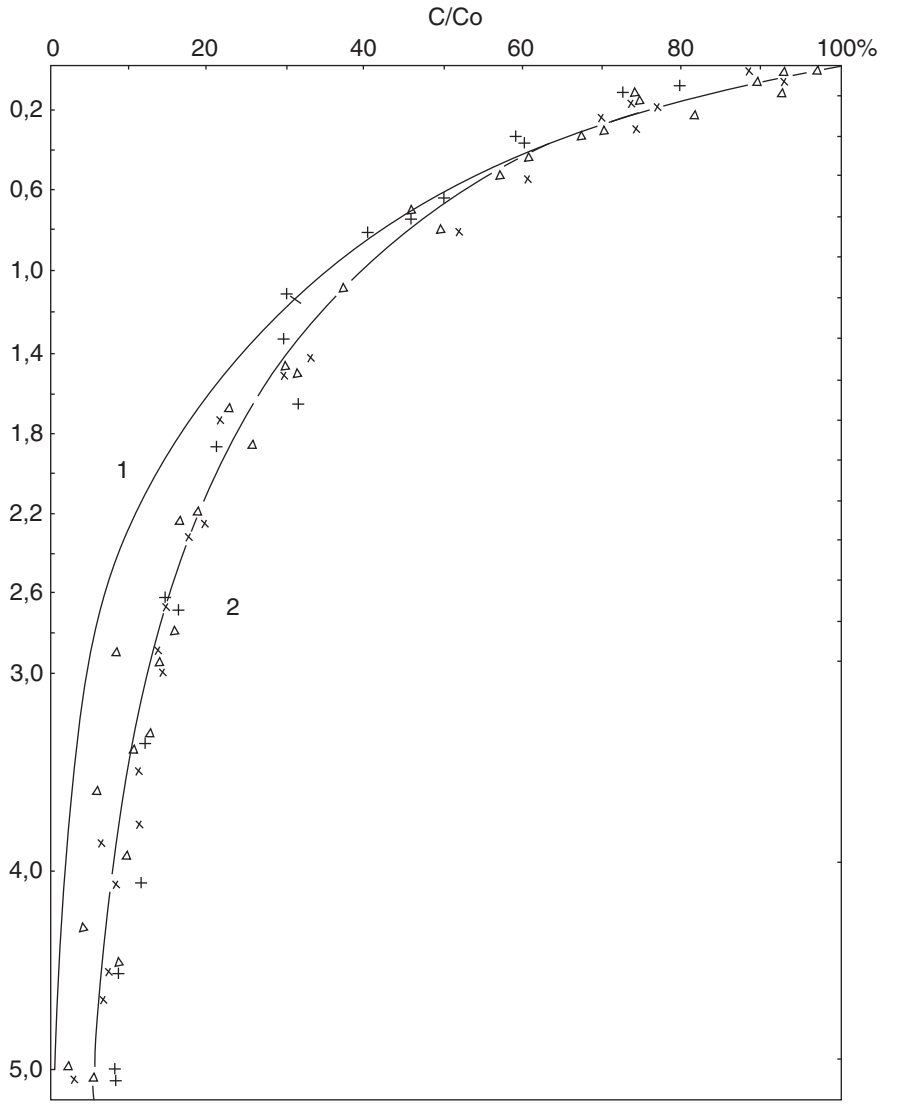


Fig. 10.1. Filtration of homogenous (1) and non-homogenous (2) suspension in porous medium—dependence of relative concentration of suspension C/C_0 on the filter depth x.

decreases with decreasing size of particles of filtrated suspension, however, this dependence has not been up-to-date quantitatively evaluated and it is only involved implicitly in the K_a value.

2.5. Single Particle Sedimentation

For the individual sphere, the substitution of the sphere diameter D for the characteristic length L and relative velocity of sphere with respect to liquid v_s for characteristic velocity v

in all dimensionless parameters is used, so that

$$Re = \rho v_s D/\mu \tag{14}$$

$$Ar = (\rho_s - \rho)\rho g D^3/\mu^2 \tag{15}$$

For the drag force acting on the sphere by laminar flow, the known Stokes equation apply:

$$F = 3\pi\mu D v_s \tag{16}$$

Because the characteristic projected area for the sphere is

$$A = \pi(D/2)^2 \tag{17}$$

the substitution of F and A into the drag coefficient gives for this case

$$f_d = 24\mu/\rho v_s D = 24/Re \tag{18}$$

Generally, f_d for the sphere is a function of Re, and it had been established by number of measurements in the range of *Re* values. The Stokes equation is valid for sufficiently small values of *Re*. The theoretically calculated first order correction is known as Oseen equation and has the form

$$f_d = (24/Re)(1 + 3Re/16) \tag{19}$$

The last term in second bracket can be used for the estimation of error when the Stokes equation is used. The Oseen equation itself can be well used for $Re < 2$.

If the drag force is attributable to gravity, then for the sphere buoyant in the liquid

$$F = (\rho_s - \rho)g(4/3)\pi(D/2)^3 \tag{20}$$

and the substitution into the drag coefficient gives for this case

$$f_d = (4/3)(\rho_s - \rho)gD/\rho {v_s}^2 = (4/3)Ar/Re^2 \tag{21}$$

The drag force is a vector and therefore having two drags a vector addition has to be applied. However, if the gravity and the flow drag acts only in opposite directions, in steady state the both drag coefficient should be equal, which gives

$$Ar = (18Re)(1 + 3Re/16) \tag{22}$$

or, solving this equation for *Re* and retaining only first order terms

$$Re = (Ar/18)(1 - Ar/96) \tag{23}$$

This equation can be used for $Ar < 20$. Substituting for dimensionless parameters, we receive for the relative velocity of sphere and liquid

$$v_s = \left((\rho_s - \rho)gD^2/18\mu\right)(1 - (\rho_s - \rho)\rho g D^3/96\mu^2\right) \tag{24}$$

In case the liquid is motionless, v_s is the velocity of steady sedimentation, in case the liquid is moving up with velocity v_{liq}, then for $v_{liq} < v_s$ the sphere is falling down, for $v_{liq} > v_s$ the

sphere is carried up. In case of nonspherical particle, all above equations hold with substitution of $D\Phi$ instead of D.

2.6. Turbulent Flow

In turbulent flow, the vector of immediate velocity **v** in a given place incessantly changes its size and direction. However, for steady flow, **v** can be divided to two parts: a mean value and fluctuation

$$\mathbf{v} = \mathbf{v}_{\mathbf{mean}} + \mathbf{v}_{\mathbf{fluct}} \tag{25}$$

where the mean (i.e., the time integral over sufficiently great time) of $\mathbf{v}_{\mathbf{fluct}}$ is zero vector (i.e., it is zero in any direction). The close fluctuations are mutually dependent, while sufficiently distant fluctuations are mutually independent.

The simplest form of turbulence is so called isotropic turbulence where characteristics of fluctuation at any point are independent of direction. The characteristics of isotropic turbulence can be exactly expressed (9) and the results can be qualitatively used for general explanation of turbulence. The intensity of fluctuation can be expressed by its root-mean-square value, and it will be signed v'. The mutual dependence of fluctuations in two points express then the so called correlation function, which is the time integral of a normalized scalar product of both fluctuations over sufficiently long time. It has value 1 for zero distance, <1 at greater distances and at sufficiently great distance it has value 0. From it can be derived important simple parameters characterizing the turbulence: the dissipation scale λ (size of small vortex) and the integral scale Λ (size of big vortex). In scales small with respect to λ, the local viscous shear gradient produced by turbulence is proportional to v'/λ. The energy dissipated in unit volume at unit time by turbulence is

$$\varepsilon' = 15(\mu/\rho)(v'^2/\lambda^2) \tag{26}$$

The decay of homogenous turbulence in its last phase proceeds so that

$$\lambda = (8\mu t/\rho)^{1/2} \tag{27}$$

$$v'^2 \approx (\mu t/\rho)^{-5/2} \tag{28}$$

where t is the time.

2.7. Coagulation

The term coagulation means the formation of aggregate particles from smaller particles by mutual contact of particles owing to their relative movement, with the following link-up of contacted particles. In case the relative movement of particles is due to thermal motion (Brownian motion), the coagulation is called perikinetic, in case the relative movement is due the movement of liquid or in liquid under external force, the coagulation is called orthokinetic. For all types of coagulation the quantitative results have been derived (10). Note that suspension with particles, which are able to form fixed mutual connections and so form greater aggregates, is called flocculating suspension. Example of nonflocculating suspension is the suspension of sand particles, examples of flocculating suspensions are many chemical

precipitates and biological activated sludge. The term flocculating is derived from the fact, that great aggregates resemble the snow flocs and so they are also called flocs.

For the rate of collisions of one particle with other particles, for perikinetic coagulation it has been derived

$$J_{\mathrm{p}} = (3/8)(kT/\mu)cf(D_1/D_2) \tag{29}$$

where k is the Boltzman constant, T temperature in Kelvin and c the concentration of particles, and D_1 and D_2 are diameters of colliding particles. The function $\mathrm{f}(D_1/D_2)$ has for equal diameters value 1 and with increase of diameters difference increases. Its explicit expression is

$$f(D_1/D_2) = 1/2 + (D_1/D_2 + D_2/D_1)/4 \tag{30}$$

For the rate of collisions of one particle with other particles due to the liquid velocity shear gradient it has been derived

$$J_{\Gamma} = (1/6)(\mathrm{d}v/\mathrm{d}z)c(D_1 + D_2)^3 f(D_1/D_2) \tag{31}$$

where the velocity direction is taken as x-axis and the maximum velocity change goes along z-axis, so that the velocity shear gradient is $\mathrm{d}v/\mathrm{d}z$. The function $f(D_1/D_2)$ has for equal diameters value 1 and with increase of diameters difference decreases. As mentioned above, for turbulence in scale small with respect to λ, the velocity shear gradient is proportional to v'/λ.

The ratio $J_{\Gamma}/J_{\mathrm{p}}$ is proportional to D^3, so that for particles sufficiently small the coagulation proceeds by perikinetic mechanism, and there exists certain critical diameter of particles over which the orthokinetic mechanism is decisive. In practical conditions this critical diameter is in range between 1 and 10 μm.

2.8. Hydrodynamic Disintegration of Aggregates

The flocs formed by coagulation of flocculating suspension are fragile and they can be therefore disintegrated by hydrodynamic forces. For qualitative insight to this problem the results of the study of breaking emulsion drops by turbulent viscous shear [11] can be used. For USBF it is important, where the size of floc is small with respect to λ. For this case, the critical diameter over which the disintegration occurs is

$$D_{\mathrm{crit}} = (K_{\mathrm{s}}/\mu)(\lambda/v') \tag{32}$$

where K_{s} is a constant depending on the nature of suspension and involving, e.g., the strength of links between particles in aggregate. Note that the hydrodynamic breakdown of greater flocs leads with great probability to two flocs with comparable diameters.

2.9. Fluidization in Cylindrical Column

The fluidization in constant and only vertical flow, represented by the fluidization in vertical cylindrical column, is the simplest case, and it was therefore the subject of thorough research (12), results of which can be then applied to more complicated cases. Note that only

fluidization in liquid–solid systems will be discussed and the phenomena pertaining to other systems (gas–solid, gas–liquid–solid) will be omitted.

The behavior of layers of granular porous medium in a vertical cylindrical column under the influence of the liquid upflow is demonstrated in Fig. 10.2. In further discussion we will assume that the column diameter is sufficiently great with respect to the particle diameter, so that the effect of column diameter can be neglected. As long as the drag force due to the flow is smaller then the medium weight, the medium is fixed. The Eq. (9) can be therefore be used giving straight ascending line in log Δp— log V coordinates, where Δp is the pressure difference over the whole height of porous medium layer, H_p, and V is apparent velocity. On other side, in fluidized state, the weight of fluidized layer buoyant in the liquid is balanced by the pressure loss, so that

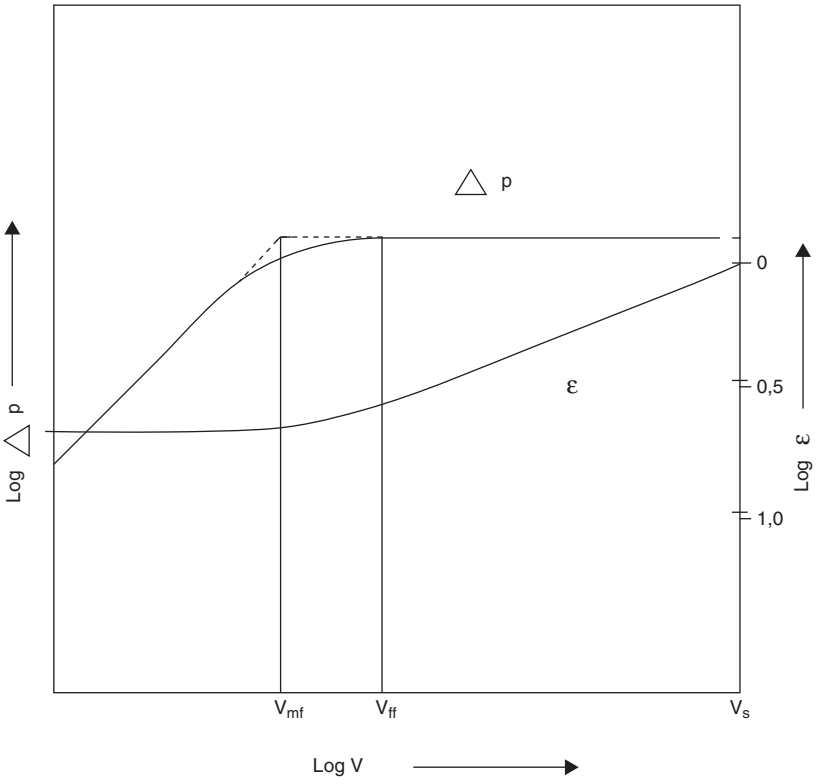


Fig. 10.2. Transition of fixed granular porous medium to fluidized layer—dependence of overall pressure difference Δp and porosity ε on apparent liquid velocity V in bilogarithmic coordinates.

$$\Delta p = (\rho_s - \rho) g (1 - \varepsilon) H_p \tag{33}$$

The total weight of granular layer remains constant, so that Δp is constant and H_p is indirectly proportional to $1 - \varepsilon$. In $\log \Delta p$—$\log V$ coordinates the transition of ascending line to horizontal line therefore occurs.

For the fluidized state it was further experimentally found that

$$V / v_s = \varepsilon^n \tag{34}$$

where v_s is the free sedimentation velocity of individual particle forming the granular porous medium. The power coefficient n depends on the value of Ar for individual particle, ranging from 4.65 for small Ar values up to 2.4 for great Ar values. Equation (34) has evidently the correct limit, because for $\varepsilon = 1$ is $V = v_{liq}$, and for $v_s = v_{liq}$ the individual particle will go neither down nor up. Note that V is there the mean apparent velocity in the sense of mean velocity as defined above for turbulence.

The Eq. (34) gives in $\log \varepsilon - \log V$ coordinates the straight ascending line. On other side, the ε value for fixed porous medium is constant, which gives a horizontal line in $\log \varepsilon - \log V$ coordinates. Graphs $\log \Delta p - \log V$ and $\log \varepsilon - \log V$ are therefore correlated—ascending line in one corresponds to horizontal line in the other and vice versa. In the transition region, both Δp and ε are changing, corresponding to the situation where, because of the drag force the particles begin to change their orientation and packing so that the porosity increases, eventually due to turbulent fluctuations the temporary local fluidization occurs, but full fluidization still does not occur. The cross section of extrapolated linear branches in $\log \Delta p$—$\log V$ graph defines the minimum fluidizing velocity V_{mf} while start of deviation from horizontal line is taken as the minimum velocity of full fluidization V_{ff}. To those velocities correspond the porosities ε_{mf} and ε_{ff}, respectively. The maximum fluidizing velocity is limited by the condition $V < v_s$.

Fluidized layer substantially influences the flow of the liquid. On one side, because it is a porous medium, it has the tendency to smooth the uneven distribution of the flow of the liquid. In case the fluidized layer has the height sufficiently great in comparison with the size of the flow nonuniformities at bottom, the mean flow of liquid at the top of fluidized layer will be uniform. On other side, the individual particles in the fluidized layer are steadily moving in all directions, and this movement resembles the thermal movement of molecules in gas. Therefore, it can be described in terms of diffusion and/or in terms of turbulence. Fluidized layer so depresses on the one side the external flow non-uniformity and turbulence but on other side it creates its own turbulence.

The fluidized layer forms on the top the horizontal surface. Its formation and stability can be simply understood from the fact that over it $v_{liq} = V < v_s$, so that the particle moved by perturbation above this level will drop down into the layer. Because the liquid flow at this surface is evenly distributed, this mechanism works over the whole surface. Note that the flow of the liquid above this surface is in the surface proximity uniform, having only small vortexes of the size comparable of individual particle diameter. However, the Reynolds number for empty column is always so great, that at greater distance from the surface of fluidized layer starts the formation of great vortexes.

From the side of bottom, the fluidized layer is unstable, because from the same reasons as above the particle falling at the bottom out of the fluidized layer would be still fall down. Fluidized layer has to be supported at the bottom. The most often used support in case of a cylindrical column is the fixed bottom. The problem of incoming liquid distribution at this bottom has to be solved.

The bottom liquid distribution can be the source of still further instability. As long as the column diameter is small with respect to the fluidized layer height, the uneven fluid distribution at the bottom will not influence the top of the fluidized layer. However, in opposite case the pressure drop in the column can decrease, if some vertical channel in the column is empty and all granular material is in other vertical channel (or pile). In fact, for the empty channel according to Eq. (33) $(\Delta p)_{\text{fluidization}} \rightarrow 0$. The cure against this instability is the flow resistance in bottom flow distribution system so that the total flow pressure drop over the column will in the case of channeling increase. This is fulfilled if the head loss in flow distribution system is greater then the head loss over the fluidized layer. The formation of channels inside fully fluidized layer with evenly distributed flow is prevented due to the turbulent diffusion of particles, which smooth the concentration differences and works therefore against the development of channeling instability.

2.10. Fluidization in Diffuser

For USBF, the fluidization spaces with upwardly broadening diffuser shape cross section are used. Three common types used are schematically shown in Fig. 10.3. It is the simple inverted truncated cone, the longitudinal prism, and the toroidal prism, which can be also described as inverted truncated cone with inserted central cone or cylinder. The first type has as a bottom input, the orifice, the second and the third types, a slot.

The upflow has in the diffuser, along with the vertical component, the horizontal component, pointing to walls, and its apparent velocity V is upwardly decreasing due to increasing flow cross section. If the velocity of liquid at the input, V_{inp}, is greater then v_{s}, and we suppose the uniform flow distribution over the whole flow cross section, then in a certain level above

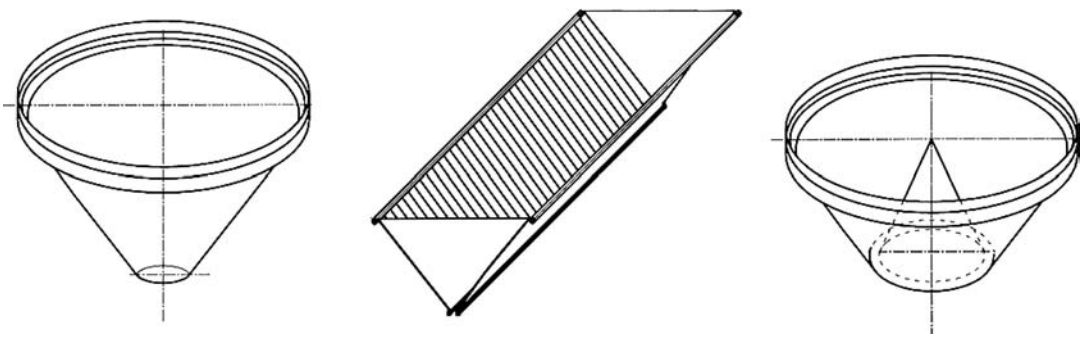


Fig. 10.3. Types of fluidization spaces commonly used for USBF.

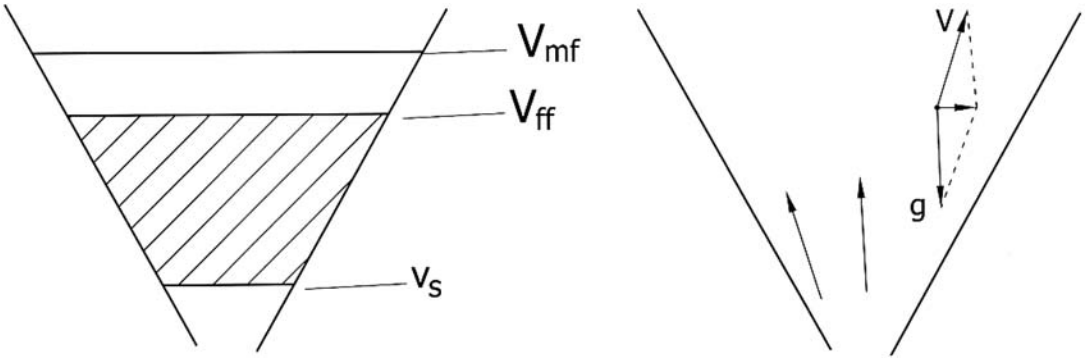


Fig. 10.4. Schema of fluidization in a diffuser.

input $V = v_s$, and in certain further levels above $V = V_{ff}$ and $V = V_{mf}$, as is schematically shown in Fig. 10.4. It is evident that the space between the first and the second levels is suitable for fluidization. The first level, $V = v_s$, would theoretically form in case of even liquid velocity distribution the base of the fluidized layer, because single particles below it would be moved up due to $V > v_s$. It will be called the base level. In fact, the liquid velocity distribution at input is nonuniform, the velocity along the diffuser axis is greater then on the sides, and the flow is there turbulent. However, having sufficient amount of granular material, the fluidized layer will evenly distribute the apparent velocity of liquid in levels in sufficient distance above the base level. Around the base level proceeds the intensive turbulent mixing gradually damping the energy and turbulence of incoming liquid and distributing the liquid flow evenly over the whole flow cross section. Due to this mixing, we can find fluidized particles even below the base level, and for keeping the granular material in a fluidized layer it is necessary that $V_{inp} \gg v_s$. However, increase of V_{inp} increases the energy and turbulence of incoming liquid, so that for obtaining fair fluidization there exists a certain optimum range of V_{inp} values. It is also evident that there exists certain minimum amount of granular material necessary for formation of fluidized layer sufficient for input energy dissipation and uniform distribution of fluid flow.

It is further evident that fluidization requires sufficient slope of diffuser walls. With too small slope, the sediments from particles will be formed on the walls, leaving along the center only free diffuser with walls formed by settled granular particles. Theoretically, for identical spheres lying on horizontal bottom, the slopes of such self-made walls with most dense (hexagonal) packing and with less dense (cube) packing will be 70.5° and 45°, respectively. Experiments with USBF proved that for smooth diffuser walls, the slope of 52° is sufficient for prevention of sedimentation of biological sludge on walls.

As has been mentioned above, evenly distributed liquid velocity apparently has also the horizontal component due to widening of diffuser. This component forces the particles to move to the diffuser walls. Against this unilateral movement acts the turbulent diffusion, nevertheless, the result is an increase of particle concentration in the vicinity of walls. The certain volume

of fluidized layer behaves, in many respects, as a dense liquid and for some purposes it can be so described. The density of such volume increases with increasing concentration of particles. The increase of particle concentration at diffuser walls therefore results in density currents flowing down along the walls. These density currents are moving against increasing liquid velocity and are therefore gradually diluted and swamped by intensive turbulent mixing in lower levels. Flow of this density currents forms vortexes contributing to the own turbulence of fluidized layer. Note that density currents do not form the even flow but they have the stochastic character—observed at given place they accidentally appear and disappear.

Because the apparent velocity upwardly decreases, it would, according to Eq. (34), correspond to upwardly decreasing porosity and therefore upwardly increasing particle concentration and upwardly increasing density of the fluidized layer. Such stratification is unstable like the layer of more dense fluid over less dense fluid. This instability creates vertical mixing, balancing the particles concentration at different levels, so that the particles concentrations are higher in bottom levels and lower in top levels then what corresponds to apparent velocity. The density currents at walls described above substantially contribute to this concentration equalization.

The greater the height of the fluidized layer for a given diffuser and given flow, the greater is the concentration of particles in fluidized layer due to decrease of V in top layers and described concentration equalization. The question as to what happens if the height goes over V_{ff} level arises. The porosity for full fluidization cannot go under $\varepsilon_{\mathrm{ff}}$, to which corresponds the upper limit of fluidized particle concentration and fluidized layer density. With increase of fluidized layer height over V_{ff} level, expansion of this limiting concentration to lower level therefore occurs. Together with it increases the intensity of wall density currents from top layers. Due to the concentration effect at the walls, the concentration inside those density currents can be even greater and will correspond to porosity between $\varepsilon_{\mathrm{mf}}$ and $\varepsilon_{\mathrm{ff}}$. The intensity of those density currents can be such that they go under the v_{s} level.

2.11. Upflow Sludge Blanket Filtration

The fluidized bed filtration, known in water treatment under the trademark name Upflow Sludge Blanket Filtration (USBF), is one of the techniques for separation of suspended solids (SS) from liquid. The other industrially used SS separation techniques are sedimentation, flotation, and fixed bed filtration. Every technique has its own field of application where its performance gives the advantage—e.g., sedimentation is suitable for separation of gravel and sand, for removal of traces of SS the fixed bed filtration is a technique of choice. USBF is superior for separation of flocculating suspension in the range of middle and high SS concentrations.

For a better understanding of complex causality in the USBF process, the schema of it is shown in Fig. 10.5. The sludge blanket is the fluidized layer of flocs, those flocs being formed from a matter of separated SS. Inside of sludge blanket there is a dynamic equilibrium—on one side the agglomeration of smaller flocs leads to increase of flocs, on other side the greater flocs are disintegrated by hydrodynamic forces resulting from the local turbulence. The result of the two counteracting processes is a certain mean floc diameter and the floc size distribution

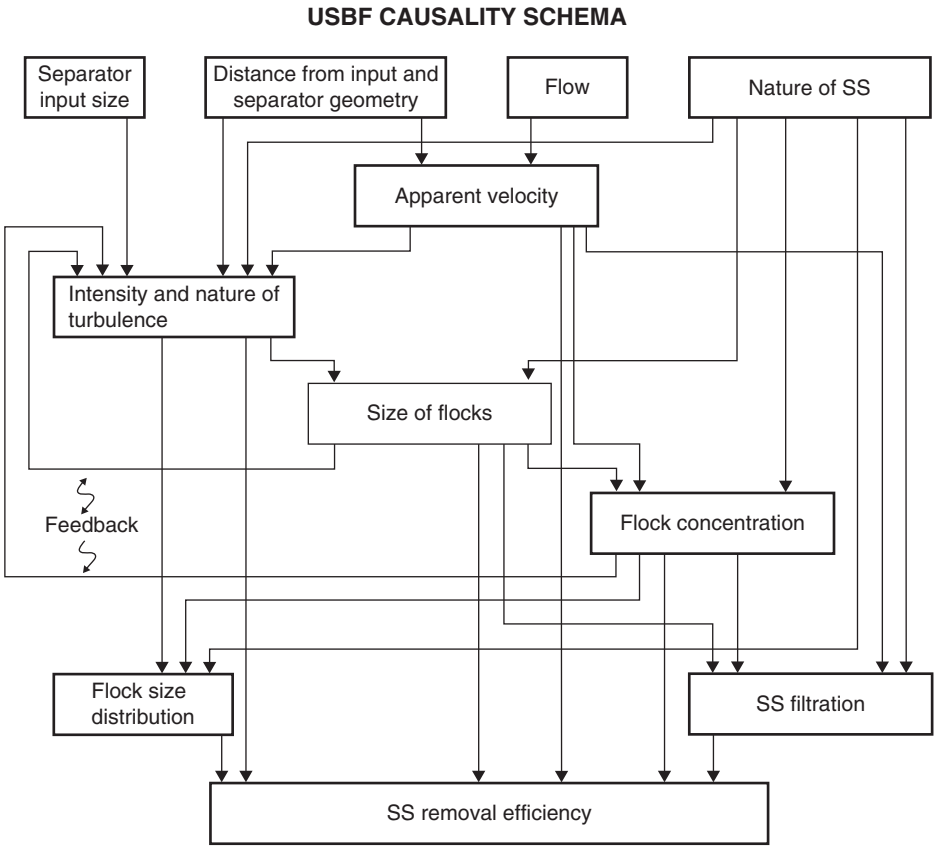


Fig. 10.5. USBF causality schema.

in a given place. Because the hydrodynamic breakdown results in flocs of comparable size and, with respect to high flocs concentration, the aggregation velocity of those flocs is high, it can be expected that the floc size distribution function will be narrow and that the great majority of flocs will have the diameter near around local mean diameter, D_{mean}, in the interval between $(1/2)D_{\text{mean}}$ and $2D_{\text{mean}}$. From Eqs. (31) and (32) follows that D_{mean} depends on the SS nature and on the local turbulence, and it increases with the decreasing intensity and with the decay of turbulence, as indicated by Eqs. (26) to (27) and (31). On the flocs diameter depends the flocs concentration—from Eqs. (24) and (34) it is seen it increases with increasing floc diameter and density and decreases with increasing apparent velocity. The apparent velocity itself depends on the geometrical factors—the separator geometry and the distance of a given place from separator input—and on the liquid inflow. The intensity and nature of turbulence at a given place has two components—one from the turbulence of incoming liquid and the second formed by its own fluidized layer turbulence, as has been analyzed in preceding section. The first component depends on geometrical factors, on initial intensity and nature of incoming liquid turbulence and on this turbulence damping, which depends on flocs concentration in a

sludge blanket. The initial incoming liquid turbulence is mainly the function of separator input size and geometry and of flow rate. The fluidized layer turbulence has been also analyzed earlier and it depends on floc concentration, floc size, and the nature of SS forming the floc. It can be seen that the interconnections between the turbulence, floc size, and flocs concentration form feedback loops. Another feedback, which cannot be expressed in the present schema, is the spatial feedback due to vertical mixing in a fluidized layer. Due to it, not only the spatial distribution of parameters but also of processes take place, and the local parameters are not the function of the local conditions but of conditions and processes in the whole sludge blanket. The process of flocs disintegration must also be mentioned, which proceeds mainly in bottom levels, while the process of flocs aggregation proceeds mainly in top levels of sludge blanket.

Suspended solids in the liquid entering the sludge blanket have, as a rule, substantially smaller size then flocs forming a sludge blanket. Its capture in sludge blanket is therefore better described in terms of filtration than of coagulation. It penetrates into the sludge blanket and the decrease of its concentration resembles the filtration curve in a fixed bed. It has been experimentally observed that the suspension from only perikinetic coagulation penetrates through sludge blanket, so that for effective capture of the suspension it is necessary to increase the suspension particles by orthokinetic coagulation. Note that with respect to the increase of the filtration coefficient with the decrease of apparent velocity, as results from Eqs. (10) to (13), the main part of the filtration proceeds in the top layers of the sludge blanket.

Suspended-solids removal efficiency depends first of all on the results of incoming suspension filtration. However, with high flow rates, the turbulent fluctuations at the surface of the fluidized layer can lift some small flocs from the fluidized layer over the adjacent layer of uniform flow so that they can then be lifted by great eddies in pure water. In such a case the removal efficiency depends also on the flocs size and floc-size distribution and on turbulence around the fluidized layer surface.

3. PRINCIPLES OF INTEGRATED USBF REACTORS DESIGN

The fluidized bed filtration is finding an ever-increasing spectrum of application in water treatment. During the last 50 years of USBF reactor design three distinct modifications of the USBF has been developed. Every one of these modifications has substantially enlarged the innovation radius of USBF to a new field of applications, which are now covering nearly the whole spectrum of the chemical and biological treatment of water.

3.1. Types of Sludge Blanket

Removal of suspended solids from liquid flowing into the sludge blanket needs, in steady state, the removal of an equivalent amount of solids from the sludge blanket to keep the total amount of solids in the sludge blanket constant. This removal proceeds in the form of an excess flocks removal. According to the means of excess floc removal, three types of sludge blanket can be distinguished: fully fluidized, partially fluidized, and combined sludge blankets. The first two types are already “classical,” the third one is a brand new development.

In the following analysis, two parameters, common in water treatment practice for the quantitative evaluation of separation process performance, will be used: surface load and

solids loading. The first describes hydraulic performance and expresses the flow of treated water coming through the unit separation surface and it is commonly expressed in $m^3/h/m^2$ units (it can be also interpreted as a velocity of water at the separation surface, because $m^3/h/m^2 = m/h$). The second describes the separation performance and expresses the flow of separated suspended solids (flux flow) per unit of separation surface and it is commonly expressed in kg $SS/h/m^2$ units.

In a fully fluidized sludge blanket, excess flocs are removed from the surface of the sludge blanket (1). Because the fluidized layer behaves as a liquid with a higher density than water, the excess of the fluidized layer overflows over upper edge of the diffuser wall into the extra space. The concentration of flocs in this overflow current corresponds to that in the top of the sludge blanket and the extra space has been used for a further increase of the sludge concentration by sludge thickening and is therefore called the thickening space. However, the spontaneous overflow creates countercurrents, which lift some flocs into the pure water zone, and thus deteriorates the sludge blanket separation efficiency. To suppress this effect, the forced flow in the overflow region has been introduced by sucking water from the thickening space (2). The hydraulic load of the sludge blanket can be very high, however, on account of a decrease in concentration of flocs in the overflow. Thus, this type of sludge blanket is suitable for separation of diluted suspensions, the concentration of which is substantially lower than the concentration of solids in the overflow. The low concentration of separated suspension results in low attainable solids loading.

In the partially fluidized sludge blanket, the top of the sludge blanket operates at water velocity below the lower limit for full fluidization. Thus, the density currents flow along separator walls, as described in the section about fluidization in diffuser. The excess flocs are removed at the bottom of the sludge blanket using the density currents, which are propagating below the sludge blanket. At the beginning, the simple return of separated suspended solids through the input was used (3). However, it was realized that the increased performance can be obtained by forced withdrawal of separated solids from below the propagated density currents (4). Because the concentration of flocs in the density currents is higher then what corresponds to the full fluidization limit, this type of sludge blanket is suitable for separation of concentrated suspensions. To obtain sufficient intensity of the density currents propagated below the sludge blanket, the velocity at the sludge blanket surface should be substantially lower then the lower limit for full fluidization. Thus, the attainable surface load for this type is low whereas the solids loading is high. Note the basic phenomenological difference of this type in comparison with fully fluidized type: for this type, the sludge blanket surface level is changing with the variation of flow and the concentration of incoming suspension and it is given by the equilibrium of processes in the sludge blanket, while in the fully fluidized type the surface level of the sludge blanket is constant and it is given by the overflow edge.

In the combined sludge blanket both preceding types are combined in such a way that their advantages are enhanced and disadvantages suppressed (5). The bottom part of the combined sludge blanket works as fully fluidized type whereas the upper part works as partially fluidized type. The fully fluidized bottom part distributes the water and solids into the upper partially fluidized part and the excess flocs are withdrawn from the density currents at the walls from

Table 10.1
Characteristics of different types of sludge blanket

Fully fluidized sludge blanket
• Very high hydraulic performance—surface load typically 4 to 5 $m^3/h/m^2$
• Low separation performance—solids loading typically up to 0.5 kg $SS/h/m^2$
• Separated sludge removal from the top
• Suitable for separation of diluted suspensions (typically below 0.2 kg SS/m^3)
• Sludge removal flow typically 10% to 20% of effluent flow
Partially fluidized sludge blanket
• Low hydraulic performance—surface load typically 0.8 to 1 $m^3/h/m^2$
• High separation performance—solids loading typically around 5 kg $SS/h/m^2$
• Separated sludge removal from the bottom
• Suitable for separation of concentrated suspensions (typically 4 to 6 kg SS/m^3)
• Sludge removal flow typically 200% to 250% of effluent flow
Combined sludge blanket
• High hydraulic performance—surface load typically 1.6 to 2 $m^3/h/m^2$
• Very high separation performance—solids loading typically around 10 kg $SS/h/m^2$
• Separated sludge removal from the side
• Suitable for separation of concentrated suspensions (typically 4 to 6 kg SS/m^3)
• Sludge removal flow typically 200% to 250% of effluent flow

the side at the middle part of the sludge blanket. Because of the fact that withdrawn density currents do not flow against liquid flow in the region of high apparent velocity, the hydraulic load can be, in comparison with partially fluidized sludge blanket, substantially higher.

The characteristics of the previous described types of sludge blanket are summarized and quantified in Table 10.1. Note that mentioned typical hydraulic performances for partially fluidized and combined sludge blankets regards to sludge from extended aeration with denitrification.

3.2. Water Treatment Systems with USBF

The above characteristics predetermine the possibility of USBF applications in water treatment systems. In those systems, the USBF is combined with processes, which eliminates the unwanted substances from the water and transforms at least part of them into the form of flocculating suspension. Processes of this type involve the majority of chemical and biological water treatment processes. A suitable shape of USBF separation space gives the possibility of integration with the advantages of all the processes in the compact reactor design.

The fully fluidized sludge blanket is suitable for the separation of diluted suspensions. Such suspensions are formed in chemical treatment, where the coagulants are added to transform the colloidal impurities present in water into a suspension. This treatment is commonly used for production of potable and utility water from surface water and it was this application,

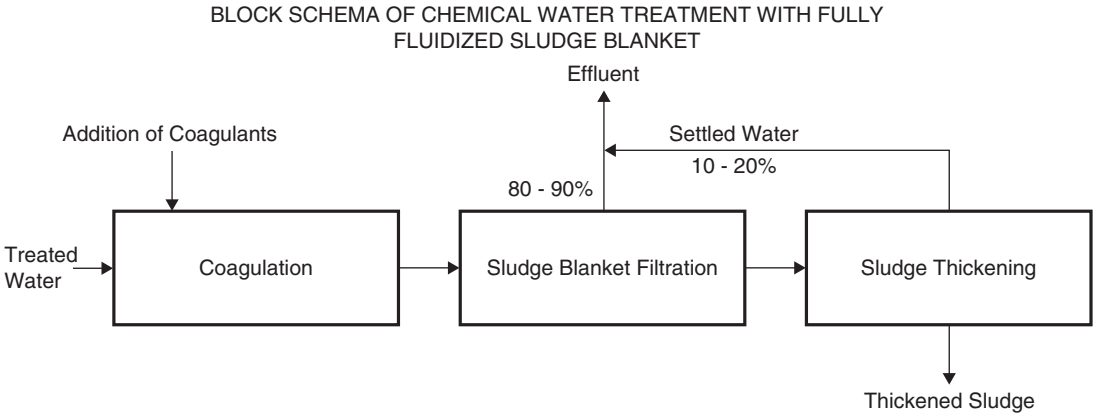


Fig. 10.6. Block schema of chemical water treatment with fully fluidized sludge blanket.

which started the development of USBF (1). The block schema of the whole process is shown in Fig. 10.6. After the addition of coagulant into the treated water stream the coagulation in it proceeds, and it needs a certain time and corresponding space. Sufficient perikinetic coagulation usually requires 5 minutes, however, addition of orthokinetic coagulation adds a further 5 to 10 minutes to obtain suspension, which is removable by filtration in fully fluidized sludge blanket. The stream of excess flocs from USBF is further concentrated by gravitational thickening, and the settled water released from it is added to the stream of effluent from USBF. The integrated chemical reactor therefore contains, along with the USBF separation space, the coagulation space and the sludge thickening space. The thickened sludge from thickening space is periodically discharged with high velocity and under the high headloss to overcome the sludge thixotropic stupor. The discharged sludge has usually 15 to 30 kg SS/m^3.

The partially fluidized and combined sludge blankets are suited for the biological wastewater treatment and their development is actually connected with biological activation process (3–5). Biological processes presently used in connection with USBF are aerobic activation, aerobic sludge stabilization, nitrification, denitrification, phosphorus removal, and selector action. Principles of their operation can be found in textbooks (13–15). Note only that for those processes the important parameter is the sludge load, and that in United States it is usually related to mixed liquor volatile sludge solids (13) and expressed in kg BOD/kg VSS/day units (or lb BOD/lb VSS/day giving the same value), while in Europe it is usually related to mixed liquor total sludge solids (16) and expressed in kg BOD/kg SS/day units (the index 5 for BOD is for simplicity omitted). For temperate climatic zone, the limit for nitrification is around 0.2 kg BOD/kg VSS/day, whereas for aerobic stabilization it is around 0.1 kg BOD/kg VSS/day.

The activation process using the suspended activated sludge needs a high concentration of activated sludge, obtainable only by sludge recirculation. In fact, the production of activated sludge from impurities in treated water represents, usually, only a few percent of present activated sludge. The separated sludge removal flow lends itself for recirculation in the

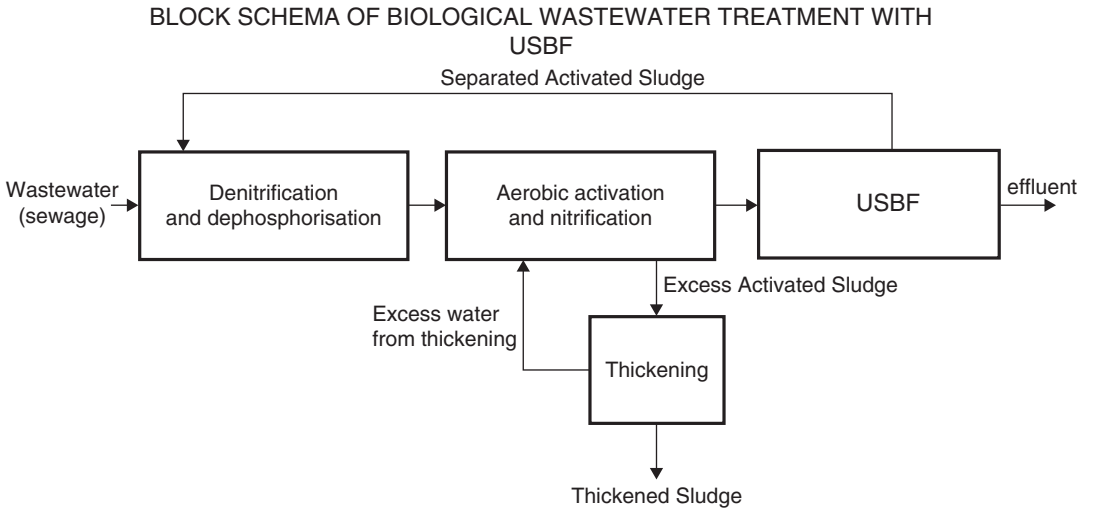


Fig. 10.7. Block schema of biological wastewater treatment with USBF.

treatment system, and the high recirculation rate can be advantageous for biological system with aerobic activation/nitrification and preceding denitrification. The block schema of the whole process is in Fig. 10.7. The integrated biological reactor (IBR) contains, along with the USBF separation space, the denitrification space and the aerobic activation/nitrification space (because those spaces are in IBR interconnected, the term zones is often used). The flow of separated activated sludge from USBF to denitrification, nitrification, and back to USBF forms the internal flow loop in IBR. The loop flow is a multiple of the flow through IBR (e.g., with sludge removal flow 200% and day and hour flow nonuniformity coefficients 1.5 and 2, the loop flow is six times the mean flow through IBR). The high concentration of activated sludge (4 to 6 kg SS/m^3, in special cases even higher) gives the possibility to use low-loaded activation process, usually with simultaneous aerobic sludge stabilization. The wastewater (in a majority of cases the sewage) after mechanical pretreatment first enters the denitrification zone, where the anoxic conditions are maintained. In the presence of nitrates brought by the internal circulation loop, there proceeds the biodegradation of organic substances from wastewater with the use of oxygen from nitrates. This will result in denitrification releasing gaseous nitrogen from nitrates. The mixed liquor then passes to nitrification zone, where the aerobic conditions are maintained. There proceeds the biodegradation of remaining organic substances and oxidation of nitrogen from ammonium and biodegraded organic compounds to nitrates. The mixed liquor from the nitrification zone enters USBF, where the activated sludge is separated and the effluent outflow equals the wastewater inflow. The mean concentration of nitrates in the effluent is in a given arrangement reduced by factor $1/(n_r + 1)$, where n_r is the ratio of loop flow to mean flow (in above example $n_r = 6$ so that the effluent will contain $100/(6 + 1) = 14.3\%$ of produced nitrates and the produced nitrates removal efficiency is therefore $100 - 14.3 = 85.7\%$; the total nitrogen removal efficiency of the whole process is

even higher because part of the incoming nitrogen is removed in wasted excess sludge; it, of course, applies for well working denitrification, particularly its sufficient volume and low oxygen concentration in it are important). The inclusion of denitrification within the process loop has still other beneficial effects. It augments pH recovery after pH decrease owing to nitrification. Suitable denitrification arrangements can create, in certain parts, anaerobic conditions with resulting increase of biological dephosphorization.

During biological treatment, the suspended activated sludge in IBR internal flow loop is repeatedly exposed to oxic and anoxic conditions in nitrification and denitrification processes. This combined with the low sludge loading and the biological selector action (screened wastewater first enters anoxic zone), results in the formation of very specific biocenosis in activated sludge. The product is activated sludge with a low sludge volume index (for sewage typically around 100 mL/g). The low sludge volume index then gives the possibility to keep high sludge concentration in activation process.

With respect to the high flow rate in the internal flow loop, the water level differences in different spaces of IBR have to be minimal to minimize the energy loss. The loop flow is generated by pumping the separated sludge from USBF to denitrification. The excess of biological sludge formed in biological processes is wasted from nitrification space. This can be built-in the excess sludge thickener decreasing the volume of wasted sludge. The proper handling of wasted sludge is important for phosphorus removal results, because the anaerobic conditions during excess sludge handling will release the captured phosphorus.

The efficiency of the USBF technology effluent is high. For typical municipal sewage and after the plant biological "run-up," mean values are better than 10/15 mg/L in BOD_5/TSS, and the residual ammonium nitrogen is well bellow 2 mg/L. The removal efficiency of total nitrogen is over 70% and the residual nitrogen is in the form of nitrates. In case of special situations, even higher removal efficiency of nitrates and TSS as well as the removal of phosphorous can be achieved. In case the tertiary treatment by microscreen filtration is added, mean values are better than 5/5 mg/L in BOD_5/TSS (for Mill Springs WWTP (Mill Bay, BC, Canada) with granular bed filtration, there was during 2002 to 2004 operation in more than 50% TSS below 1 mg/L).

4. EXAMPLES OF USBF INTEGRATED TREATMENT REACTORS IMPLEMENTATION

Implementation of the USBF technology can be divided into three applications. The first is represented by chemical and the other two by biological treatment. The differences between them have resulted in specific design of water treatment facilities for chemical and biological treatment. While in the case of chemical modification the original USBF principle has been preserved during the whole period of implementation, in the case of biological modification a new impulse in the form of COMBI USBF has provoked an important new breakthrough in the whole technology of biological treatment.

4.1. Chemical USBF Integrated Reactors

The first large municipal plant for chemical treatment with USBF technology was built in 1955 for the city of Brno (Czech republic) with the capacity of 1,000 L/s for potable water. Schematic cross section of the USBF reactor used is presented in the Fig. 10.8, and an inside view of the plants eight reactors is in the Fig. 10.9. The reactor has a concentric configuration with sludge blanket compartment of toroidal prism shape. The central cone is used as mixed coagulation space with a treated water inlet at the top and the water flows from it into the sludge blanket through bottom slot. The outer space around sludge blanket compartment serves as a sludge thickening space. The pure water is collected by troughs above the sludge blanket and the excess sludge withdrawal proceeds from the top of the sludge blanket. The flow of sludge into the lower part of sludge thickening compartment is rectified by screening baffle, behind which proceeds a sucking of a part of the treated water through this compartment into the effluent. This effective hydraulic arrangement of the USBF chemical treatment reactor was patented in 1952 (1) and represents the first generation of USBF integrated reactors, which are used today and cover a wide range of reactor capacities up to 150 L/s.

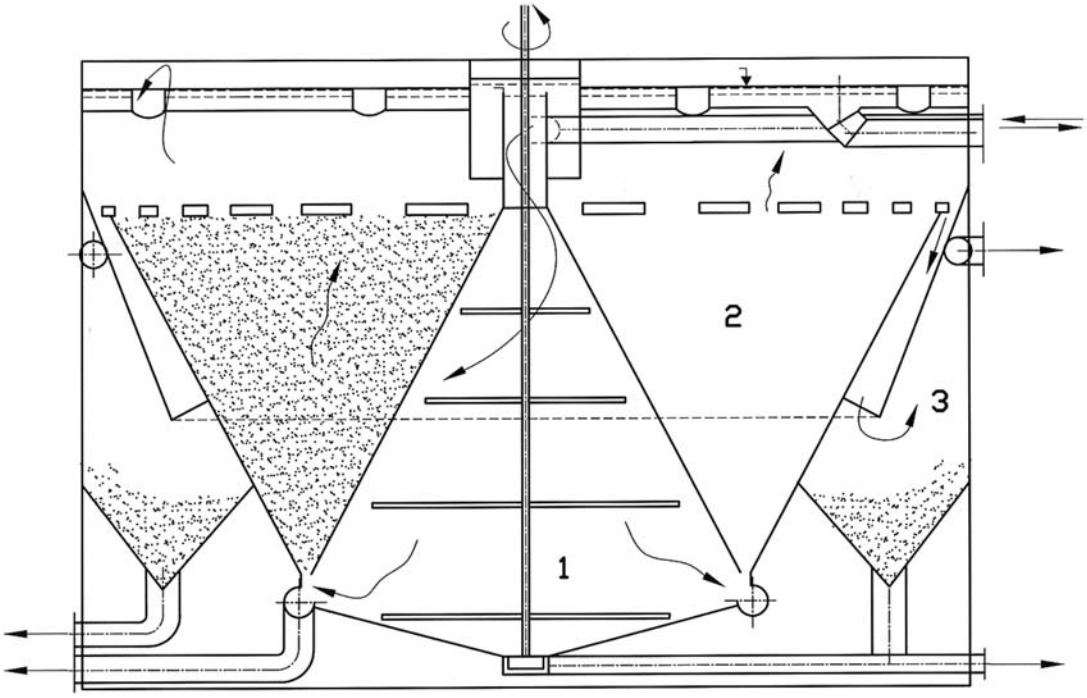


Fig. 10.8. Cross section of concentric chemical USBF reactor. 1—coagulation space, 2—USBF separation space, 3—sludge thickening space.

Fig. 10.9. Inside view of Municipal Treatment Plant for potable water with total capacity 1000 L/s (8×125 L/s). Brno, Czech Republic (1956).

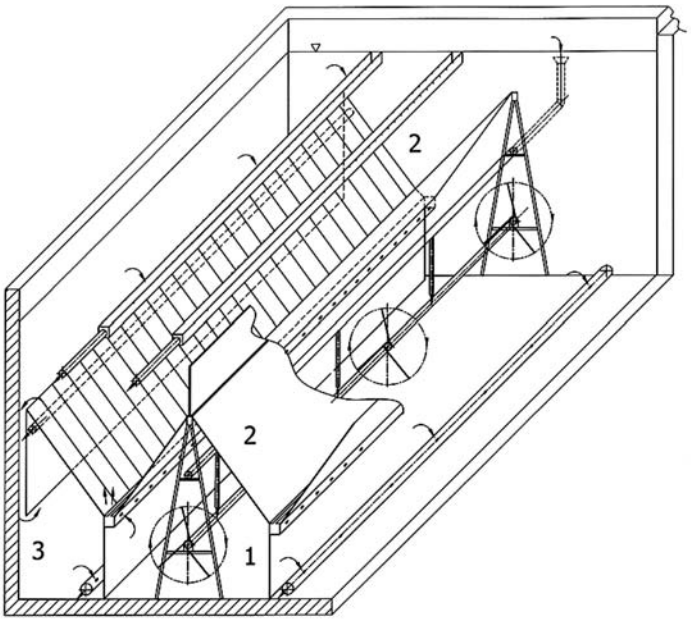


Fig. 10.10. Axonometric view of horizontal chemical USBF reactor. 1—coagulation space, 2—USBF separation space, 3—sludge thickening space.

For larger capacities a horizontal configuration of the chemical USBF reactor has been developed. A schematic cross section of such an integrated reactor is illustrated in the Fig. 10.10 and a picture of a chemical treatment plant for coal mine water with capacity $2 \times 250\,\mathrm{L/s}$ for elimination of Fe and Mn is illustrated on the Fig. 10.11. The horizontal configuration of the reactor preserves all hydraulic principles of the concentric one, and differs only in the shape of functional compartments, having prismatic sludge blanket compartments with adherent longitudinal coagulation and sludge thickening compartments. As in the concentric concept of chemical USBF reactor, a portion of the treated water with excess sludge is sucked from the surface of the sludge blanket into the sludge-thickening compartment. After the sludge gravitational thickening in this compartment, the treated water is added to the final effluent from the sludge blanket. The horizontal configuration of the integrated chemical USBF reactor covers the largest capacities in chemical treatment of water up to several cubic meters per second.

In the range of small capacity chemical treatment plants, the USBF technology for potable water can be integrated with rapid filter in one compact technological unit. An example of such complex transportable unit is shown on the Fig. 10.12, in which the granular multilayer filter is incorporated into the spherical structure of USBF chemical reactor.

4.2. First Generation of Biological USBF Integrated Reactors

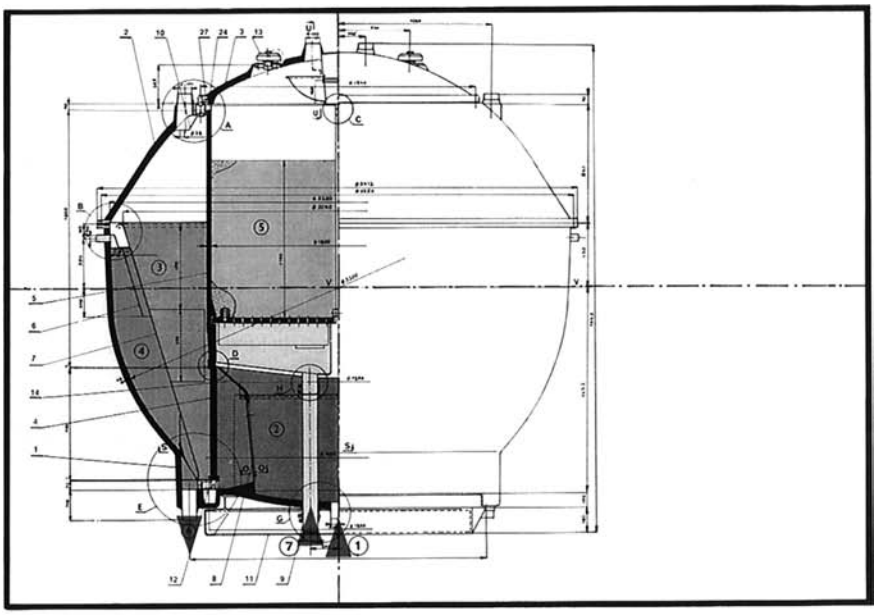
At the beginning of the application of the USBF technology to the biological wastewater treatment, the partially fluidized sludge blanket was used. The biological USBF integrated

Fig. 10.11. Picture of chemical USBF reactor 2 × 250 L/s. Coal mine Svatava, Czech Republic.

reactors of this type was using diffuser shaped sludge blanket separator having (with the exception of the smallest units) the form of an inverted truncate cone for the small units, with central entrance system at the bottom, and the form of a longitudinal prism for larger units with longitudinal slot shape entrance. For recirculation of the mixed liquor from the conical shaped USBF separator, the central collection system connected to the recirculation pump was used, and in the case of the prism shaped USBF separator a longitudinal collection pipe at the bottom of the separator equipped with a recirculation pump was provided. For excess sludge withdrawal, a sludge thickener is inserted into the activation zone of the integrated reactor. The concentrated excess sludge was pumped to the sludge storage tank and further dewatered by means of mechanical dewatering equipment as for example the Sieb belt press or decanter.

The industrial implementation of the first generation of biological USBF integrated reactors was covering the whole range of municipal, industrial, and agricultural waste water and a wide range of capacities reaching from the smallest domestic treatment plant for individual family houses to the largest communal and industrial applications. For demonstration of the scope of the application of this technology, some of the typical cases are illustrated on the following examples.

Figure 10.13 shows a domestic sewage treatment plant with the capacity of five population equivalents (PE) and in Fig. 10.14 an axonometric view of the same unit (17) (note that in this



① Raw water supply
② Reaction zone
③ Sludge blanket zone
④ Sludge concentration zone
⑤ Multilayer filter bed
⑥ Concentrated sludge outlet
⑦ Water supply for washing sand filter bed

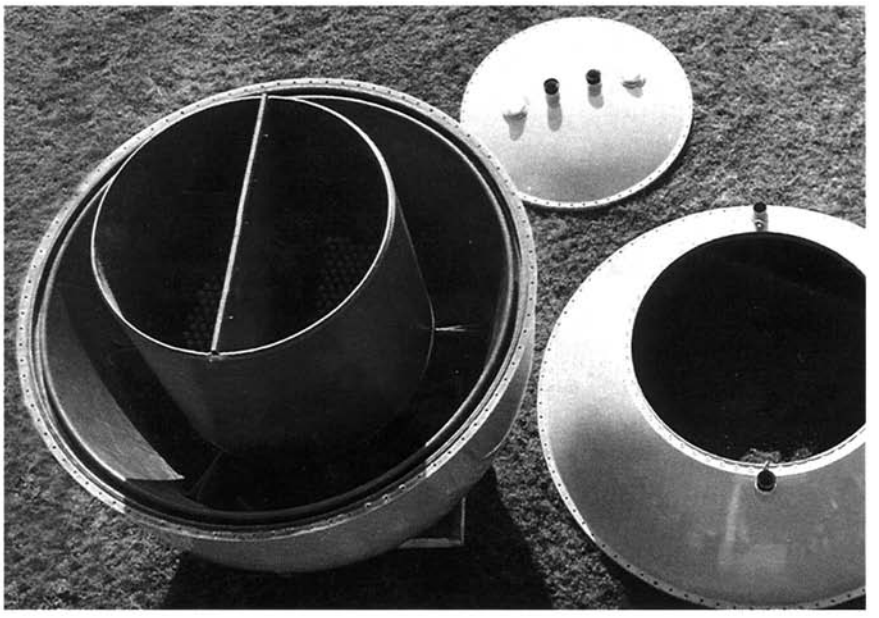


Fig. 10.12. Cross section and picture of package potable treatment plant 5 L/s.

Fig. 10.13. Domestic wastewater treatment plant with capacity five person equivalent (PE).

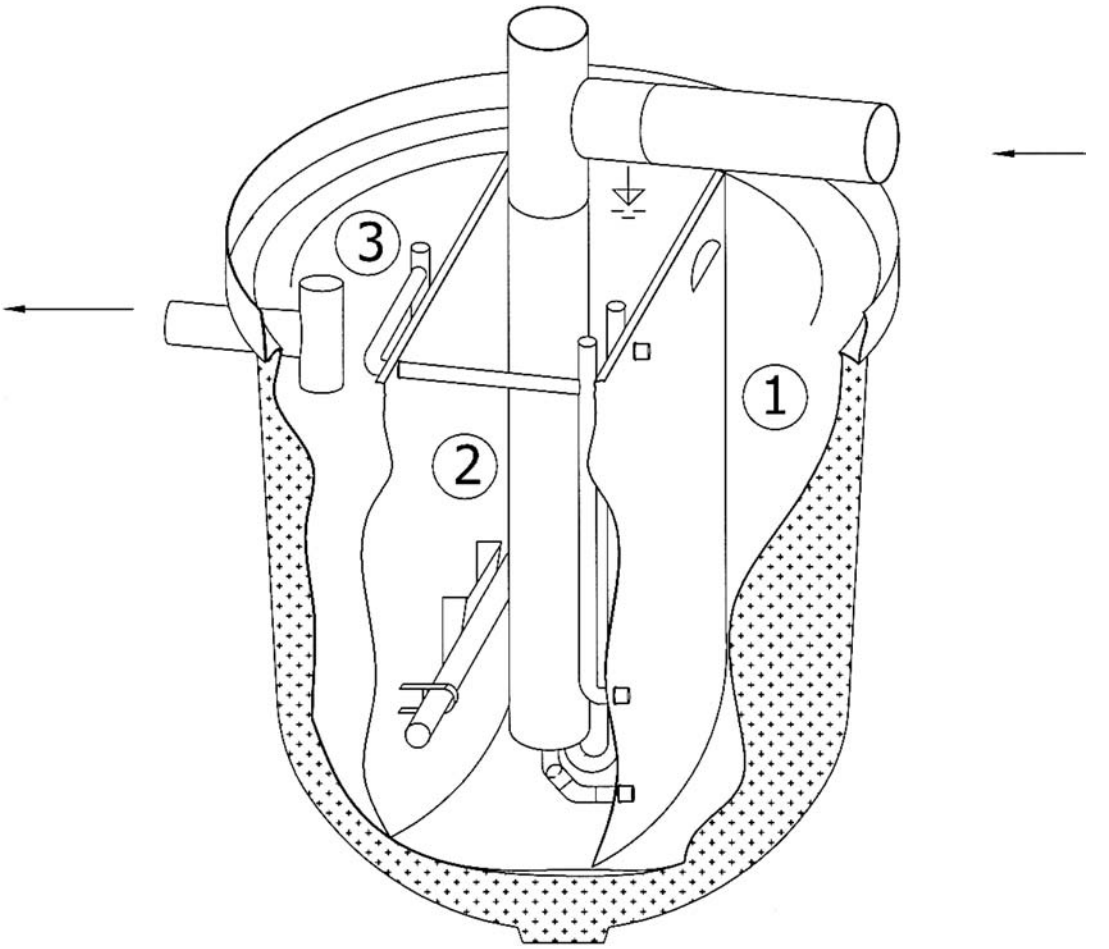


Fig. 10.14. Axonometric view of domestic WWTP with capacity 5 PE. 1—denitrification, 2—aerated activation and nitrification, 3—USBF.

type a special shape of USBF compartment is used, however, the general principles of sludge blanket operation are conserved). Figure 10.15 represents a package plant of larger capacity in the range of 50 to 500 PE transportable on a lorry. An example of municipal WWTP with 20,000 PE capacity is on Fig. 10.16 and the ground plan of the same plant is on Fig. 10.17. Schematic cross section of this integrated biological USBF reactor is given on the Fig. 10.18. Example of the industrial application of the first generation of the biological USBF reactors is shown on Fig. 10.19, representing WWTP for treatment of slaughterhouse wastewater with capacity 1,000 m^3/day, and system Agroclar for pig liquid manure treatment with capacity of 22,000 pigs is on Fig. 10.20.

Fig. 10.15. Package WWTP for 100 PE during transport.

An example of mean year operation results measurement of plants Pinzolo (Italy), with capacity 32,000 PE (population equivalent), and Hatě—free shop (Czech republic), with capacity 560 PE, both equipped with partially fluidized USBF, are given in Table 10.2 below. Added are mean values from 25 effluent measurements from different domestic plants with capacity 5 PE.

Note that there is a difference in wastewater—in Hatě there is a concentrated sewage from WC and wastewater from a restaurant, while in Pinzolo there is a combination of municipal sewage with irregular discharge of wastewater from alimentary industry (dairy, slaughterhouse). It has been observed that after the shock discharge of great amounts of industrial water, a temporary deterioration of plant purification efficiency follows.

The Pinzolo plant is noteworthy from another standpoint. This plant, located in a very aesthetically exposed area of the Alps, is housed in an architecturally attractive building (Fig. 10.21), providing an example of a successful architecture for district WWTP suitable for location directly in an urban environment.

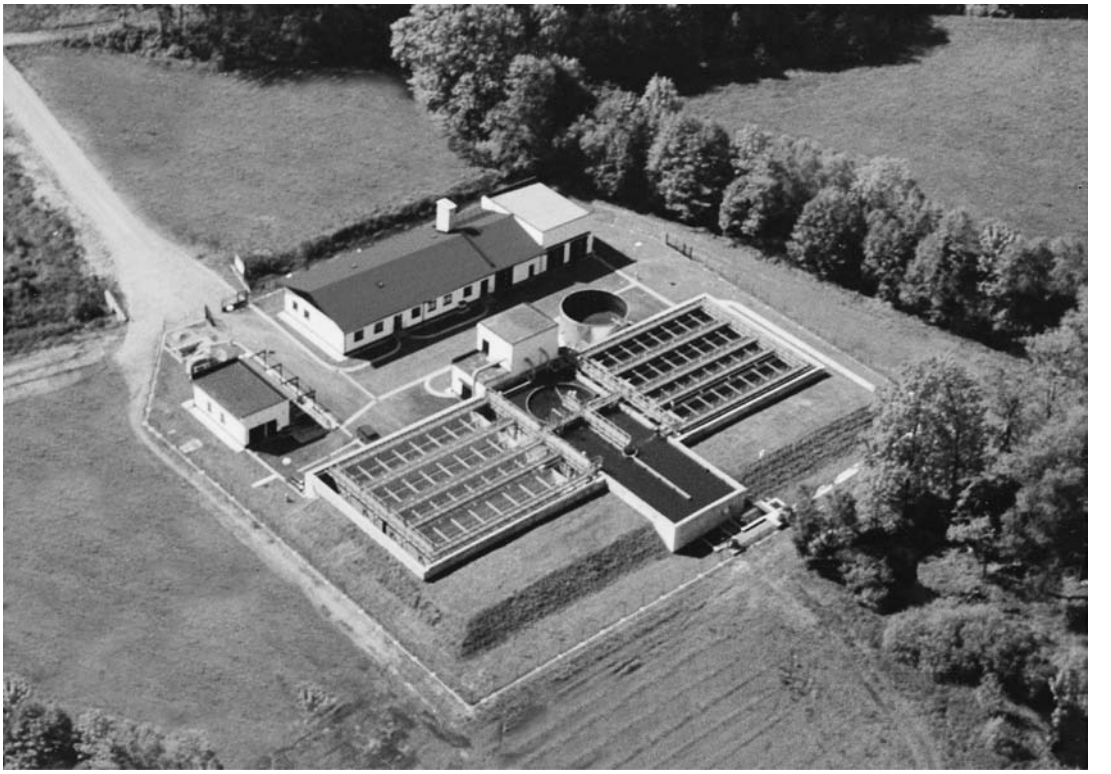


Fig. 10.16. View of a WWTP of Nové Město n. M., Czech Republic (CITYCLAR, 20,000 PE).

4.3. Second Generation of Biological USBF Integrated Reactor

Ever increasing demands for higher hydraulic capacity of USBF separation initiated the development of a new concept of integrated reactor for biological treatment using the principle of combined sludge blanket revealed above (COMBI USBF). The large increase of sludge blanket filtration efficiency of the COMBI USBF manifests itself not only in reduction of the required separation surface and volume of the separator, but plays a decisive role in influencing the geometrical configuration and the overall layout of the entire integrated reactor. The result is a completely new generation of the USBF integrated reactors (with exception of the smallest capacity units).

Examples of the USBF integrated reactor with the COMBI sludge blanket are demonstrated on the new types of OXICLAR and CITYCLAR units. For smaller capacity, cone shaped separators are used. Figure 10.22 illustrates the axonometric view of the general layout of the integrated biological reactor of this type. Figure 10.23 shows the assembled separator cone being transported for erection into a reactor. For larger capacity, a prism shaped separator is more suitable. Figure 10.24 shows an axonometric view of a biological integrated reactor with the COMBI USBF type separator, and Fig. 10.25 shows an overall picture of a large municipal WWTP in construction equipped with COMBI USBF separators. Separators in one

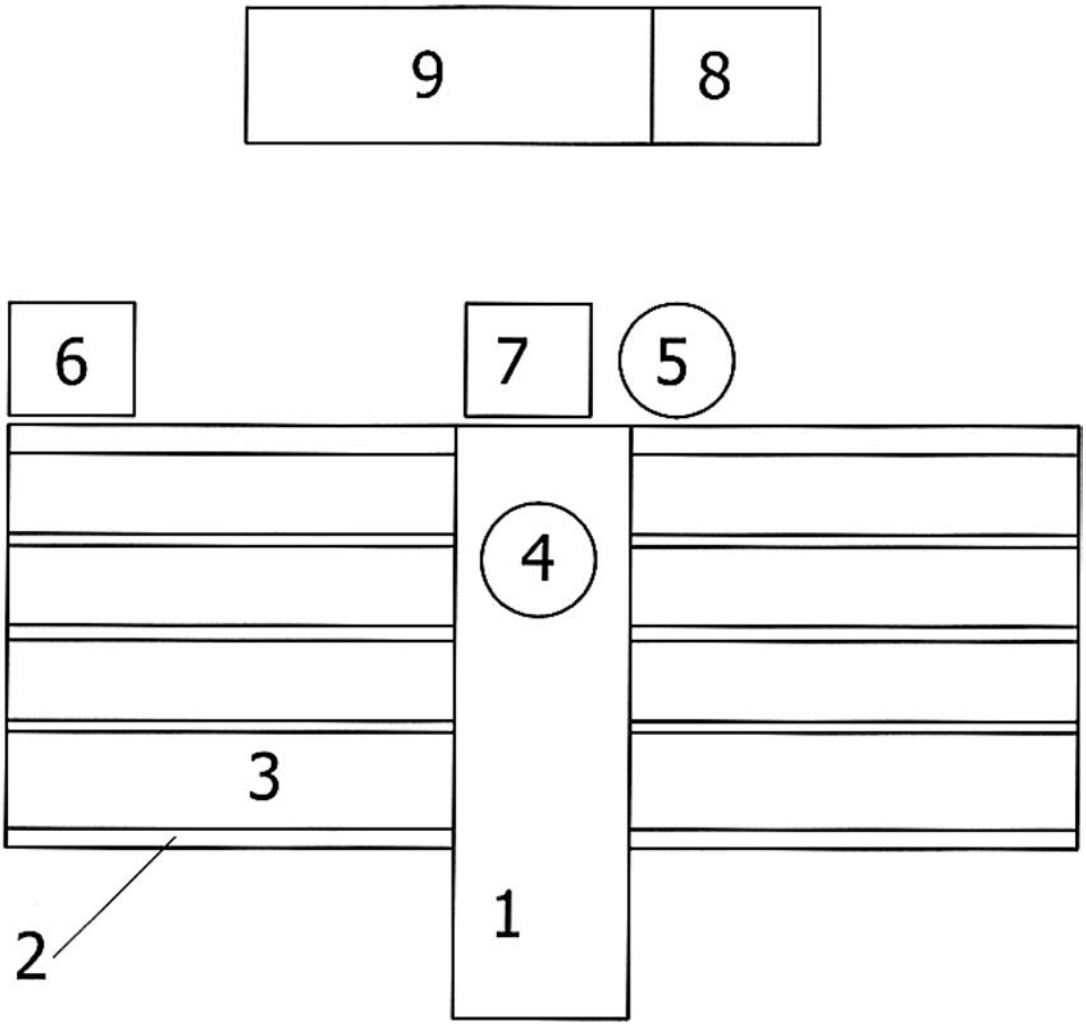


Fig. 10.17. Schematic disposition of Nové Město n. M. WWTP. 1—denitrification, 2—aerated activation and nitrification, 3—USBF separators, 4—sludge thickener, 5—storage of thickened sludge, 6—mechanical pretreatment, 7—blowers, 8—sludge dewatering, 9—office.

part of the municipal WWTP Pinzolo (Italy), mentioned above, has been reconstructed in 2001 for COMBI USBF to increase the hydraulic capacity of the plant for elimination the rainwater bypass. It can serve as an example of plant upgrading using the COMBI USBF. The picture of the reconstructed part is shown in Fig. 10.26 and results of its operation are given in Table 10.3. Noteworthy is the effluent total suspended solids (TSS) decrease in comparison with Table 10.2 results. This effect has been also visually observed as effluent transparency difference for reconstructed and unreconstructed WWTP Pinzolo parts and it has since been observed in many other COMBI USBF installations.

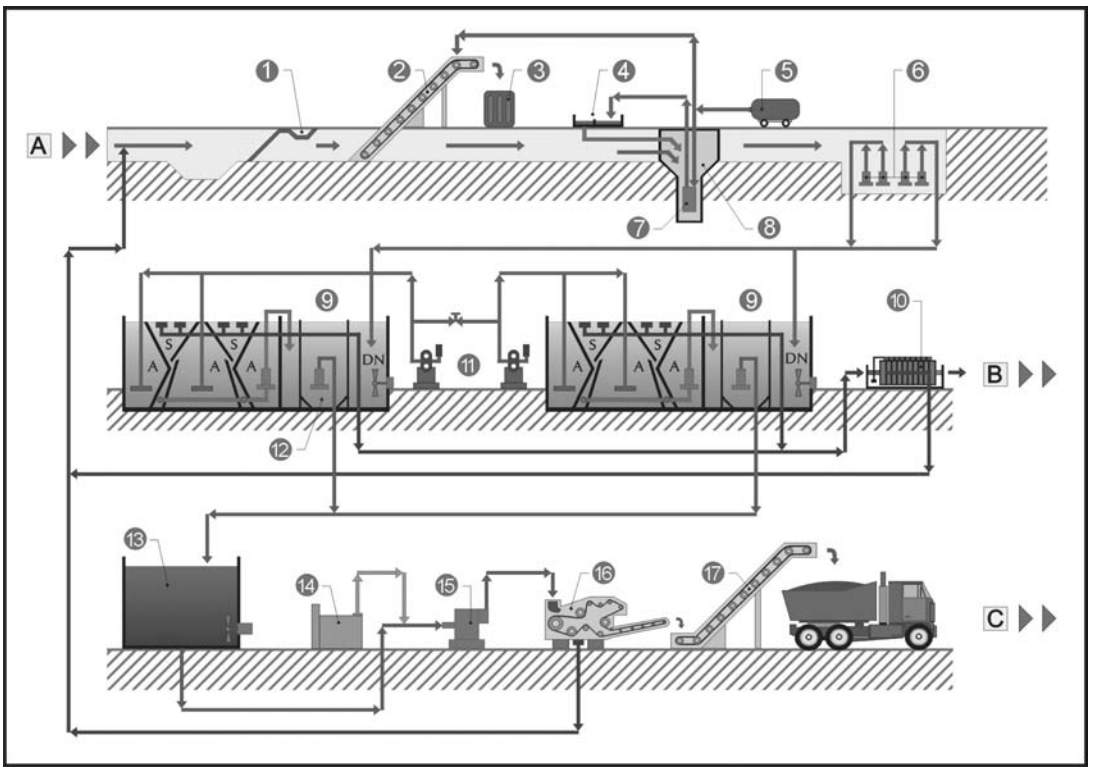


Fig. 10.18. Schematic flow diagram of USBF CITYCLAR type used in Nové Město n. M. WWTP. 1—course screens, 2—fine self cleaning screens, 3—debris container, 4—sand dewatering, 5—compressor, 6—sewage pumps, 7—air lift pump, 8—sand trap, 9—biological reactors, 10—microscreen drum filter, 11—blowers, 12—sludge thickener, 13—sludge storage tank, 14—chemical aids, 15—sludge pump, 16—belt press, |A|—wastewater inflow, |B|—treated effluent, |C|—stabilized dry sludge, A—aerated activation and nitrification, DN—denitrification, S—USBF separators.

5. ADVANCED WASTEWATER TREATMENT SYSTEMS

The greatly enhanced capacity of the COMBI USBF technology can be exploited advantageously in three innovations:

- Increasing the hydraulic capacity of the USBF reactor
- Lowering the sludge load by increasing the activated sludge amount in USBF reactor to reach the higher purification effect
- Minimization of the USBF clarifier dimensions and increasing the compactness of the integrated reactor design

From these major innovations three major fields of perspective applications result:

- Upgrading of existing WWTP
- Decentralization of sewerage systems using individual small WWTP
- Waste water reclamation

Fig. 10.19. View of a slaughter house WWTP Pro Sus, Italy (1000 m^3/day).

All these applications represent solutions of global problems for which the existing state of art of treatment technology is insufficient, not capable to offer adequate solution.

5.1. Upgrading of Conventional Municipal WWTP

Conventional municipal wastewater purification plants are predominantly designed for operation in a medium sludge load regime (0.3 to 0.4 kg BOD/kg VSS/day). They mainly remove easily biodegradable organic matter expressed in BOD_5, while the substantial part of nitrogen pollution, particularly in its toxic ammonium form, passes through. This, from modern point of view, insufficient purification loads the environment with pollution. Moreover, owing to the towns' expansion, many old plants have insufficient capacity, which further brings substantial pollution load.

Classical plant layout consists of mechanical pretreatment, primary sedimentation tanks (in smaller plants often omitted), biological activation basins and secondary sedimentation tanks. Simple and relatively inexpensive upgrading of such plants can be achieved by reconstruction of existing tanks and basins to USBF reactors by changing their connections and inserting USBF separators in it (along with this principal reconstruction some other changes like addition of mixers to denitrification, eventual exchange of aeration system for more effective one, etc. are often necessary). Such a reconstruction results in a substantial increase of

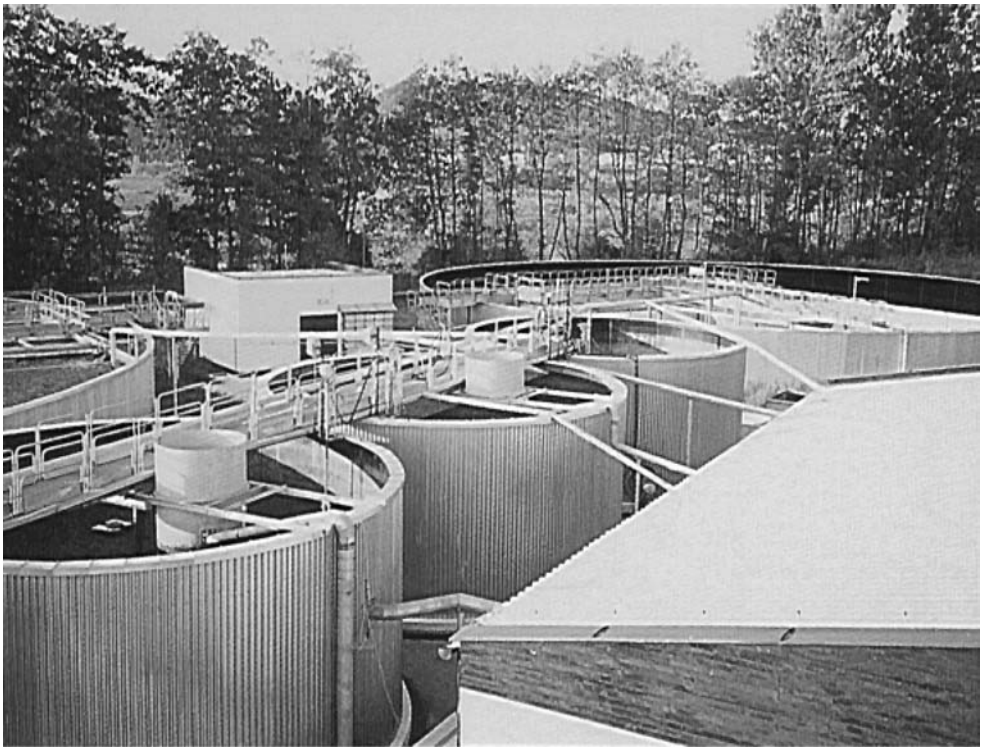


Fig. 10.20. Pig manure WWTP for 30,000 pigs, Budča, Slovak Republic.

Table 10.2
Results of operation of some plants with USBF

Plant	Year		TSS	BOD_5	COD	NH_4-N	P_{total}	NO_3-N
Pinzolo	1997	Influent (mg/L)	179	171	324	21.7	3.4	0
		Effluent (mg/L)	14	6.8	21.4	0.8	1.1	7.4
		Removal (%)	92.2	96.0	93.4	96.3	67.6	–
	1998	Influent (mg/L)	190	157	323	17.9	3.5	0.09
		Effluent (mg/L)	15	8.5	20.8	1.2	1.1	6.9
		Removal (%)	92	94.6	93.6	93.3	68.6	–
	1999	Influent (mg/L)	172	178	304	17.0	3.6	0.06
		Effluent (mg/L)	12	7.4	18.1	1.4	1.3	6.0
		Removal (%)	93.0	95.8	94.0	91.8	63.9	–
Hatě	1996	Influent (mg/L)	1132	1265	2080	77.3	–	0.22
		Effluent (mg/L)	6.6	5.6	70.7	0.25	–	2.79
		Removal (%)	99.4	99.6	96.6	99.7	–	–
Domestic	1996–1997	Effluent (mg/L)	5.9	6.1	19.5	0.57	–	6.2

Fig. 10.21. Overall picture of WWTP Pinzolo, Italy (32,000 PE).

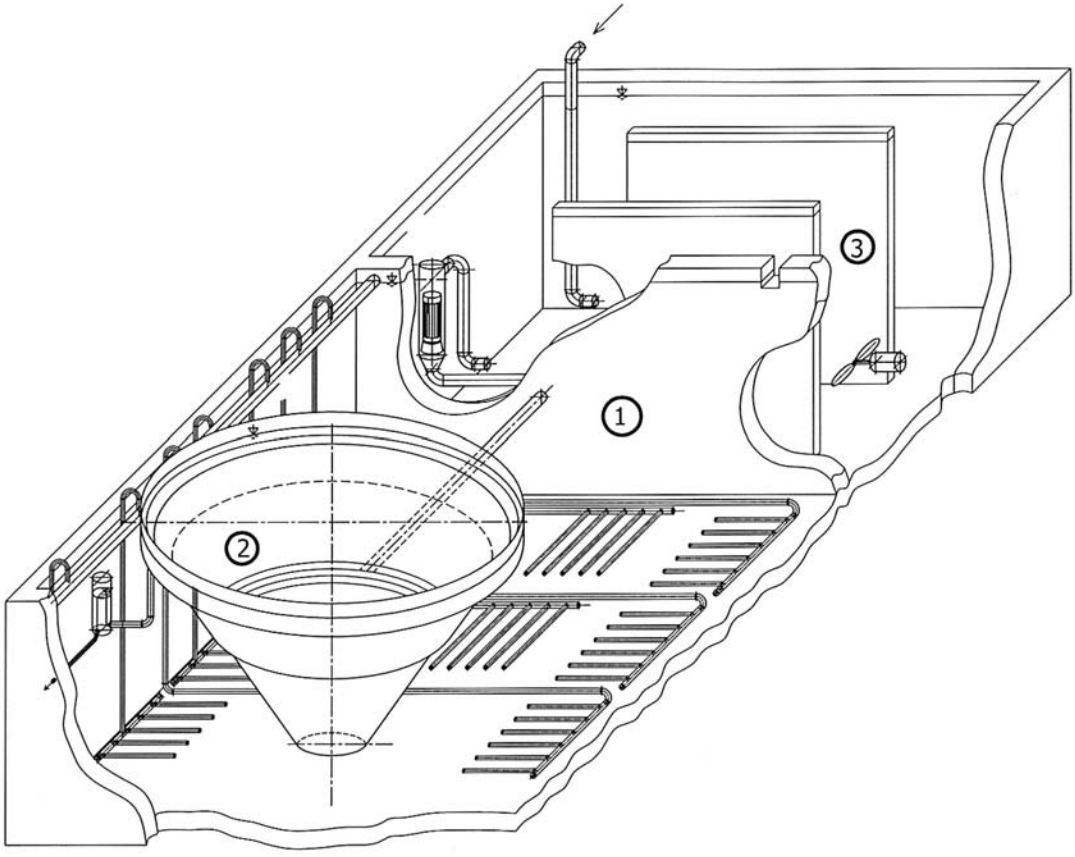


Fig. 10.22. Axonometric view of OXICLAR WWTP with COMBI USBF. 1—aerated activation and nitrification, 2—COMBI USBF separator, 3—denitrification.

Fig. 10.23. Picture of the funnel-shaped COMBI USBF separator.

activated sludge content, enabling reduction of sludge load and increase of sludge age. Thus, the advanced wastewater treatment technology with low sludge load purification regime using nitrification/denitrification can be achieved. The amount of activated sludge has to be, in comparison with conventional WWTP, increased circa two times for nitrification/denitrification and three to four times for aerobic stabilization (limits 0.2 and 0.1 kg BOD/kg VSS/day, respectively). The purification efficiency of organic matter is also increased, both in BOD_5 and COD, however, the main effect is the total removal of toxic ammonium and substantial reduction of total nitrogen pollution. Moreover, decrease of phosphorus content by biological dephosphorization will also occur. The N and P content decrease is essential, e.g. for the suppression of surface water eutrophization.

The increase of hydraulic capacity owing to COMBI USBF along with substantial increase of activated sludge content gives a possibility to increase simultaneously the plant capacity

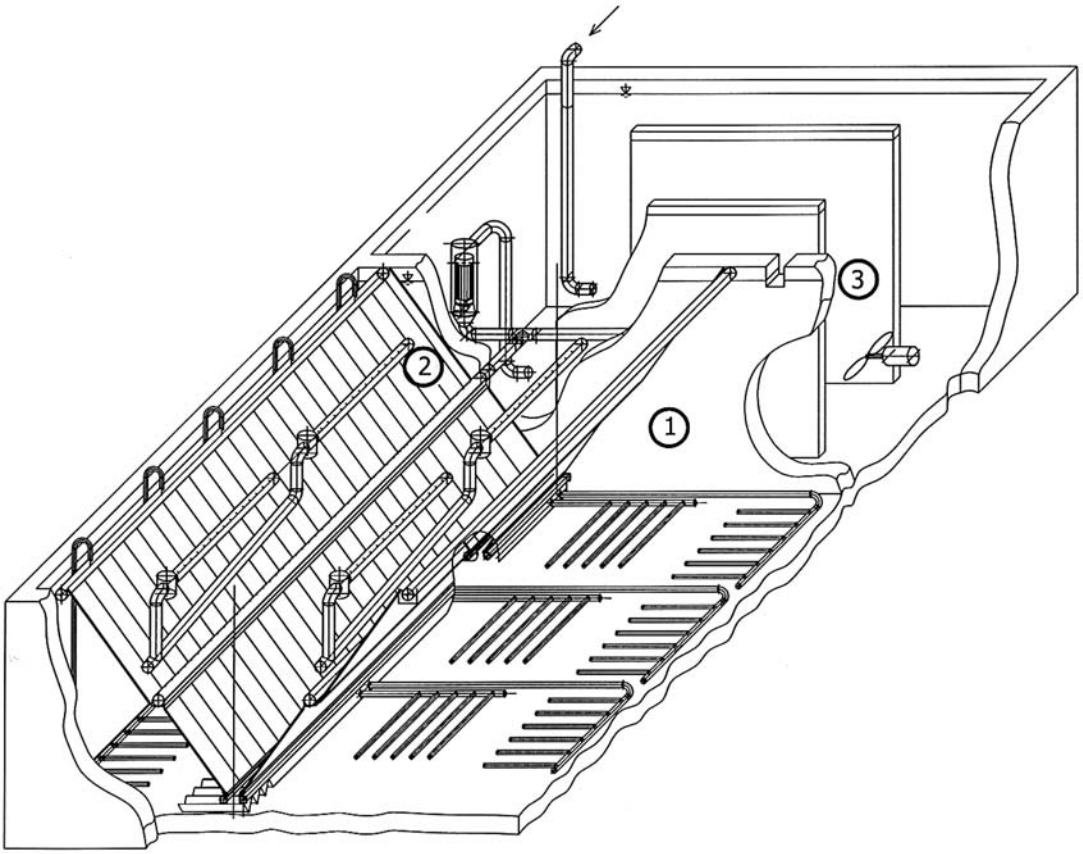


Fig. 10.24. Axonometric view of CITYCLAR WWTP with COMBI USBF. 1—aerated activation and nitrification, 2—COMBI USBF separator, 3—denitrification.

and purification efficiency. This combined increased can be solved individually according to concrete parameters of upgraded plant. The increase of the hydraulic capacity can be also exploited for elimination the by-pass of the rainwater to the recipient.

5.2. Decentralized Sewerage Systems

The expansion of centralized sewerage infrastructure out of the city center brings the sewerage investment cost to an unbearable level. This problem has led to the adaptation of alternative solutions, which can be characterized as a decentralized sewage system. The decentralized sewage system is characterized as a system with a number of local sewage purification plants. It will substantially reduce the sewage infrastructure owing to the total investment cost of sewage handling. A prerequisite for this concept is the existence of small purification plants with high purification efficiency, small dimensions, high reliability, simple operation, low service demands and acceptable investment cost. The USBF technology fulfills all those requirements. The small USBF reactors have been originally invented for small

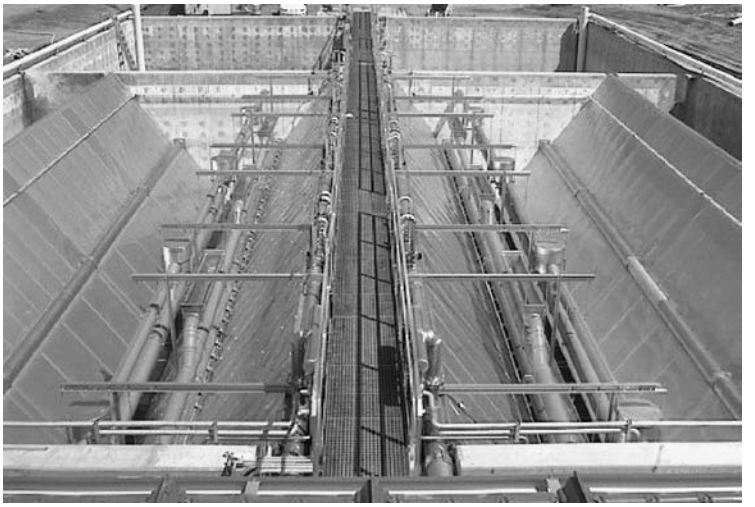


Fig. 10.25. View of WWTP with COMBI USBF in construction.

Fig. 10.26. Picture of USBF reactor in WWTP Pinzolo (Italy) after upgrading to COMBI USBF.

Table 10.3
Results of operation of WWTP Pinzolo part reconstructed to COMBI USBF, March–April 2002

		TSS	BOD_5	COD	NH_4-N	P_{total}	NO_3-N
Mean	Influent (mg/L)	206	162	361	24.6	2.55	0.19
	Effluent (mg/L)	5	5	17.6	0.48	0.67	3.45
	Removal (%)	97.6	96.9	95.1	98.0	73.7	–
Maximum	Influent (mg/L)	377	201	516	38.6	3.05	0.61
	Effluent (mg/L)	10	5	20	0.61	0.96	4.8
Minimum	Influent (mg/L)	106	92	194	16.0	2.06	0.1
	Effluent (mg/L)	1	5	13	0.34	0.52	1.3

isolated sources of sewage and industrial pollution like individual family houses, individual apartment buildings, hotels, department stores, schools, gas stations, roadhouses, camps, farms, slaughter houses, dairies, breweries, etc.

The capacity of integrated USBF biological reactors is covering the range from small domestic units for individual family houses to large units for towns and industry. Application of these integrated biological reactors brings the possibility of application of decentralized sewerage systems in rural regions, city satellites, tourist areas, new housing micro-regions, sanitation of towns and cities in developing countries, etc. The advanced purification technology used yields a purified water quality highly surpassing the quality obtained by conventional purification systems. The attainable quality is so high that it enables even the treated water reuse. The actual treated water handling of course depends on the local conditions and requirements. It can range from the direct discharge to local watercourse or to storm sewerage through infiltration into soil to water reuse as described in the next section.

A mass application of a great number of small integrated biological reactors in one area creates the possibility of simple and economical service of reactor operation. It is estimated that for domestic reactors of MICROCLAR type one serviceman can cover the service for up to 500 units in one restricted area. In fact, the first experience with such a decentralized system is in "Horná Potóň" village (Slovak republic), here, all the houses were equipped with MICROCLAR domestic units instead of the conventional sewerage system.

Another special example of advantageous district purification plant application is the sanitation of large urban conglomerations in developing countries. Large districts in these agglomerations are still without water and sewerage. In many cases the development of the infrastructure is planned or is under way. The application of district purification plants for this case can bring substantial savings in civil works leading to substantial financial savings and a decrease of disturbance of residents by civil works during the construction of sewer.

5.3. Wastewater Reclamation and Reuse

Mankind is today facing a serious shortage of fresh water. The magnitude of this crisis is demonstrated by the fact that 80 countries are now experiencing critical shortages of water. Forty percent of the world—more than two billion people—have no access to clean water or

sanitation, and the demand for water is doubling every 20 years. The supply of fresh water cannot keep pace with the needs of the growing population. It appears that "new" water will have to be mainly reclaimed from existing by better management and by water recycling and reuse. The wastewater reclamation and reuse is, in fact, the only alternative today and in the future to provide an adequate supply of water to the population and industry.

Industrial water reuse represents an area comprising a number of specific cases, each being characterized by a given wastewater pollution characteristics and the process water requirements, ranging from simple up to very complicated problems. As a simple case the water reuse from car washing can be mentioned, an example of a complicated case of water reuse can be seen in some chemical productions.

The sewage treatment for water reuse is the most often encountered case. From the point of view of water quality requirements for reuse, two categories can be there defined. The first category is for the so-called Brown-water concept, which comprises the use of water for sanitary purposes and irrigation. The second category is for the so-called Blue-water concept, which requires the resulting quality of potable water.

The common practice in sewage treatment is biological purification, enabling ecologically acceptable discharge of effluent into environment. For water reuse, the biological sewage purification should be complemented by the tertiary treatment. The tertiary treatment technology depends on the water reuse category and on the quality of effluent from the biological purification step.

The classical biological sewage treatment technology requires the sophisticated and complicated train of physicochemical processes to reach required resulting water quality. A typical example is the Orange County project in California using the complicated multiple stage tertiary treatment ending by reverse osmosis to reach the water quality for direct injection to groundwater aquifer (c.f. description in (13)). The main reason for application of very complicated and extremely expensive tertiary treatment was that the conventional biological treatment left high residual content of ammonium and organic substances.

The fundamental feature of new advanced wastewater reclamation and reuse technology is that it shifts the emphasis of treatment from physical and chemical processes to a biological process. The possibility of this technology has been opened owing to the advantage of USBF. The removal of ammonium can be achieved by low loaded activation with added nitrification. This brings the advantage for simple and effective disinfection for the less demanding applications as, e.g., the irrigation.

Biology can also be employed in the reduction of organic matter often considered "bioresistant." To achieve this, the activated sludge treatment process operating in the region of very low activated sludge loading, referred to as "superactivation," is employed.

For a long time it was prevailing in the wastewater treatment expert's community the opinion that the very low loading of activated sludge deteriorates the purification effect (see, e.g., graphs on pages 190 and 191 in Ref. (16)). However, their experience with very low loaded USBF systems revealed that reverse is true. The indication of it was obtained in "Agroclar" systems for purification of liquid pig manure, where the very high COD removal was required for effluent discharge in ecologically sensitive region. It was observed that with increasing sludge age, the COD gradually decrease, and after more than 1 year of adaptation

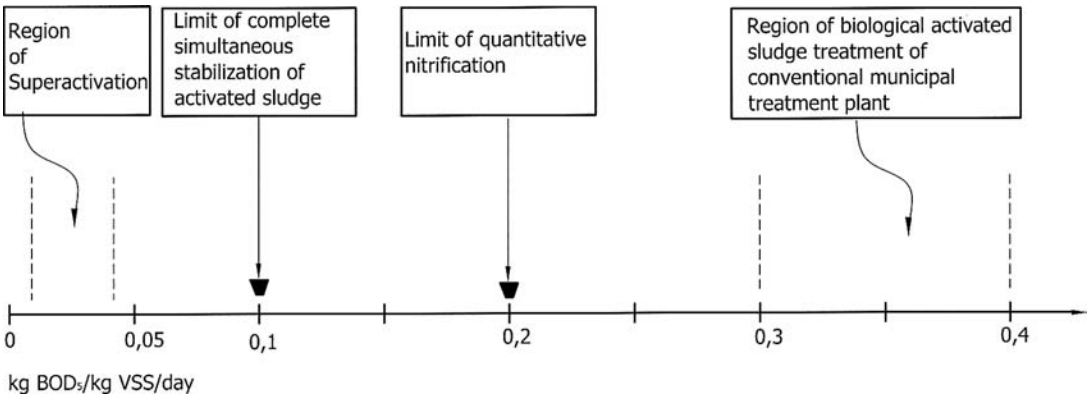


Fig. 10.27. Regions and limits of different biological activation regimes.

with very low sludge load (around 0.03 kg BOD/kg VSS/day), the COD removal exceeded 99%. Moreover, it has been observed that the COD decrease in this range was not reflected in BOD values, BOD being still in units of ppm. It indicates that there proceeds the biological process of removal of organic matter, commonly considered as "bioresistant." This process has been referred to under name "superactivation."

In fact, the superactivation is a logical extension of observed increase of COD removal with decreasing sludge load in municipal sewage treatment. For good operating medium sludge load plants, the effluent COD value usually ranges 60 to 100 mg/L, while for low loaded USBF plants, the usual effluent COD value is around 20 mg/L (see, e.g., Table 10.2).

The regions of different operation regimes are graphically illustrated on Fig. 10.27. Note that the limits of all processes are dependent on water temperature in reactor and numerical values given in Fig. 10.27 apply for temperate climatic zone; increase or decrease of water temperature by 5 °C will be reflected by multiplication or division of given values with the factor 1.4.

From Fig. 10.27 it follows that superactivation requires an order of magnitude more activated sludge within the system than the conventional activated sludge process and more than two times more in comparison to the extended aeration process. This is the main reason why superactivation has been practically beyond reach of the conventional wastewater systems. The USBF process had changed it. Operating at higher sludge concentrations and giving better reactor volume use, the quantity of sludge within the system is substantially increased (see calculated example). The resulting biological removal of ammonia and the reduction of the organic matter facilitate then the tertiary treatment. For production of utility water of near potable water standard it is sufficient to apply coagulation in granular filter bed and appropriate disinfection. Note that owing to the ammonia removal and COD reduction, the bactericidal effectiveness of disinfection is enhanced and the formation of toxic by-products is suppressed. Particularly in chlorination, the bactericidal effectiveness is enhanced by three orders of magnitude, owing to the difference in free and bounded chlorine action. In special cases with very high water quality requirements, further polishing steps can be applied such as

ozonization and active carbon adsorption for removal of last traces of refractory organics and reverse osmosis for reduction of dissolved salts. All these physicochemical processes will be economical only when treating already sufficiently purified water, such as after the superactivation process. Therefore, the COMBI USBF in combination with superactivation can play decisive role in solving one of the limiting factors of the further growth of civilization—global shortage of fresh water.

6. DESIGN EXAMPLE OF ADVANCED TREATMENT SYSTEMS

The COMBI USBF technology is providing a great flexibility for upgrading of conventional municipal WWTP. Such upgrading can be oriented either to increase of the WWTP capacity, or to increase of its treatment efficiency for effective tertiary treatment in complex wastewater reclamation systems. Both those cases are demonstrated in calculations below. However, combined solution in the form of simultaneous increase of capacity and treatment efficiency is also possible, depending on the chosen process parameters.

6.1. Upgrading of Classical Municipal WWTP

The plant using a classical technology with extended aeration located in subtropical region has the following dimensions and layout of biological treatment (real case):

- Six rectangular aeration tanks, each $84.2 \times 21.2 \times 4.75$ m, tank volume $8{,}400\,\mathrm{m}^3$ (tanks have at bottom the sides reinforcement). Total aeration volume $V_{\mathrm{aer}} = 50{,}400\,\mathrm{m}^3$.
- Six secondary circular clarifiers, each having diameter 38 m, surface $1{,}134\,\mathrm{m}^2$, volume $3{,}474\,\mathrm{m}^3$. Total clarifiers surface $6804\,\mathrm{m}^2$, volume $V_{\mathrm{cl}} = 20{,}844\,\mathrm{m}^3$.

Plant operation results (1 year average):

- Influent: Q $48{,}800\,\mathrm{m}^3$/day, COD 665 mg/L, BOD_5 350 mg/L, TSS 240 mg/L, TKN 47 mg/L, NH_3 31 mg/L
- Secondary effluent: COD 51.4 mg/L, BOD_5 6.8 mg/L, TSS 12.6 mg/L, TKN 5.5 mg/L, NH_3 3.9 mg/L
- Aeration tank mixed liquor concentration: $C_x = 4.3\,\mathrm{kg\ SS/m}^3$

Task:

Propose reconstruction with USBF technology using all existing tanks

1. To increase plant capacity keeping the same sludge age
2. To increase sludge age keeping the same capacity

Solution:

General: All existing clarifiers will be converted to preceding denitrification, all existing aeration tanks will be converted to aerated activation with nitrification and built-in longitudinal USBF separators with combined sludge blanket. A standard COMBI USBF separator has the width 6.5 m and the surface to volume ratio 0.645/m. For calculations, minimum water temperature 15°C is supposed, the activated sludge production is calculated according to Imhoff [14], flow variation and flow rate variation factors are with respect to plant size taken to be 1.2 and 1.4, respectively.

For existing plant with above parameters, calculation gives following results:
Plant BOD_5 loading TBOD $= 48{,}800 \times 0.35 = 17{,}080\,\text{kg/day}$
Plant SS loading TSS $= 48{,}800 \times 0.24 = 11{,}712\,\text{kg/day}$
Total mixed liquor suspended solids (in activation) MLSS $= 50{,}400 \times 4.3 = 216{,}720\,\text{kg SS}$
Activated sludge production
EXSS $= 0.6 \times \text{TSS} + 0.6 \times (1 - 0.072 \times t_{ts}/(1 + 0.08 \times t_{ts})) \times \text{TBOD} = 0.6 \times 11{,}712 + 0.6 \times (1 - 0.072 \times 18.4/(1 + 0.08 \times 18.4) \times 17{,}080 = 11{,}783\,\text{kg SS/day}$
Sludge age $t_{ts} = \text{MLSS/EXSS} = 216{,}720/11{,}783 = 18.4\,\text{day}$
Sludge load $B_x = \text{TBOD/MLSS} = 17{,}080/216{,}720 = 0.079\,\text{kg BOD}_5\text{/kg SS}$
Maximum flow rate $Q_{max} = (48{,}800 \times 1.2 \times 1.4)/24 = 3416\,\text{m}^3\text{/h}$
Clarifiers maximum surface load $v_s = 3416/6804 = 0.5\,\text{m}^3\text{/h/m}^2$
Clarifiers maximum solids loading $\text{SL} = v_s \times C_x = 0.5 \times 4.3 = 2.15\,\text{kg SS/h/m}^2$

Case 1.

Proposed wastewater flow $Q = 88{,}500\,\text{m}^3\text{/day}$
Maximum flow rate $Q_{max} = (88{,}500 \times 1.2 \times 1.4)/24 = 6195\,\text{m}^3\text{/h}$
Proposed USBF maximum surface load $v_s = 1.7\,\text{m}^3\text{/h/m}^2$
Required total separator surface $P_s = Q_{max}/v_s = 6{,}195/1.7 = 3644\,\text{m}^2$
Corresponding total separator volume $V_s = P_s/0.645 = 3{,}644/0.645 = 5{,}650\,\text{m}^3$
Separators length in one aeration tank $l_s = P_s/6.5/6 = 3{,}644/6.5/6 = 93.4\,\text{m} = 2 \times 46.7\,\text{m}$
Aerated activation volume $V_{AA} = V_{aer} - V_s = 50{,}400–5{,}650 = 44{,}750\,\text{m}^3$
Denitrification volume $V_{DN} = V_{cl} = 20{,}844\,\text{m}^3$
Total activation volume $V_A = V_{AA} + V_{DN} = 44{,}750 + 20{,}844\,\text{m}^3 = 65{,}594\,\text{m}^3$
Denitrification proportion $100 \times V_{DN}/V_A = 100 \times 20{,}844/65{,}594 = 32\%$
Proposed mixed liquor concentration $C_x = 6\,\text{kg SS/m}^3$
Corresponding maximum USBF solids loading $\text{SL} = v_s \times C_x = 1.7 \times 6 = 10.2\,\text{kg SS/h/m}^2$
Total mixed liquor suspended solids $\text{MLSS} = V_A \times C_x = 65{,}594 \times 6 = 393{,}564\,\text{kg SS}$
Plant BOD_5 loading $\text{TBOD} = \text{Q} \times 0.35 = 88{,}500 \times 0.35 = 30{,}975\,\text{kg/day}$
Plant SS loading $\text{TSS} = \text{Q} \times 0.24 = 88{,}500 \times 0.24 = 21{,}240\,\text{kg/day}$
Activated sludge production for sludge age 18.4 day
EXSS $= 0.6 \times 21{,}240 + 0.6 \times (1 - 0.072 \times 18.4/(1 + 0.08 \times 18.4) \times 30{,}975 = 21{,}369\,\text{kg SS/day}$
Sludge age $t_{ts} = \text{MLSS/EXSS} = 393{,}564/21{,}369 = 18.4\,\text{day}$
Sludge load $B_x = \text{TBOD/MLSS} = 30{,}975/393{,}564 = 0.079\,\text{kg BOD}_5\text{/kg SS}$

Resulting effect:

Plant capacity increase $88{,}500/48{,}800 = 1.81$ times

Case 2.

Proposed USBF maximum surface load $v_s = 1.2\,\text{m}^3\text{/h/m}^2$
Required total separator surface $P_s = Q_{max}/v_s = 3416/1.2 = 2847\,\text{m}^2$
Corresponding total separator volume $V_s = P_s/0.645 = 2847/10.645 = 4414\,\text{m}^3$

Separator's length in one aeration tank $l_s = P_s/6.5/6 = 2847/6.5/6 = 0\ 73$ m
Aerated activation volume $V_{AA} = V_{aer} - V_s = 50{,}400 - 4414 = 45{,}986\,m^3$
Total activation volume $V_A = V_{AA} + V_{DN} = 45{,}986 + 208{,}44\,m^3 = 66{,}830$
Denitrification proportion $100 \times V_{DN}/V_A = 100 \times 20{,}844/66{,}830 = 31\%$
Proposed mixed liquor concentration $C_x = 8$ kg SS/m^3
Corresponding maximum USBF solids loading $SL = v_s \times C_x = 1.2 \times 8 = 9.6$ kg SS/h/m^2
Total mixed liquor suspended solids $MLSS = V_A \times C_x = 66{,}830 \times 8 = 534{,}640$ kg SS
Activated sludge production for sludge age 54.7 day
$EXSS = 0.6 \times 11{,}712 + 0.6 \times (1 - 0.072 \times 54.7/(1 + 0.08 \times 54.7)) \times 17{,}080 =$ 9768 kg SS/day
Sludge age $t_{ts} = MLSS/EXSS = 534{,}640/9{,}768 = 54.7$ day
Sludge load $B_x = TBOD/MLSS = 17{,}080/534{,}640 = 0.032$ kg BOD_5/kg SS/day
MLSS increase $534{,}640/216{,}720 = 2.47$ times

Resulting effect:

Sludge age increase $54.7/18.4 = 2.97$ times
Sludge production decrease $9{,}768/11{,}793 = 0.83$ times

Comment:

1. In case 1 there are two separators along both long walls of aeration tank, in case 2 there is one separator along one long wall of aeration tank.
2. With respect to sub-tropical region, in case 2 the superactivation is reached, so that the substantial increase of COD removal can be expected.

NOMENCLATURE

a = filtration coefficient
A = projected area
Ar = Archimedes number
B_x = sludge load, kg BOD_5/kg SS/day
BOD = biochemical oxygen demand, mg O_2/L
c = concentration of particles
C = concentration of suspended solids (SS)
C_0 = the initial value of C
COD = chemical oxygen demand, mg O_2/L
D = Particle diameter
D_1 and D_2 = diameters of different particles
Eu = Euler number
F = drag force
f_d = drag coefficient
$f_D(D)$ = distribution function
$f(ax)$ = suspension non-homogeneity function
g = gravitation constant

H_p = height of porous medium layer
J_p = rate of particles collision in perikinetic coagulation
J_Γ = rate of particles collision due to the liquid velocity shear gradient
k = Boltzman constant
K_a = filtration constant
K_R = empirical constant
K_s = disintegration constant
L = characteristic length
Ma = dimensionless filtration criterion
n = empirical power coefficient
n_r = ratio of loop flow to mean flow
p = pressure due to the liquid flow
P = specific surface
PE = population equivalent
Q = mean flow rate, m^3/day
Q_{max} = maximum flow rate, m^3/h
Re = Reynolds number
t_{ts} = sludge age, day
T = temperature in K
TKN = total Kjeldahl nitrogen
TSS = total suspended solids
v = characteristic velocity
v_s = free sedimentation velocity, relative velocity of particle with respect to liquid
v_{sl} = clarifiers maximum surface load, $m^3/h/m^2$
v' = intensity of turbulent fluctuations
V = apparent velocity
V_{ff} = minimum velocity of full fluidization
V_{mf} = minimum fluidizing velocity
V_{inp} = velocity of liquid at the input of diffuser
Δp = pressure difference
ε = porosity
ε_{ff} = porosity corresponding to V_{ff}
ε_{mf} = porosity corresponding to V_{mf}
ε' = energy dissipated by turbulence
Φ = surface factor
λ = dissipation scale
Λ = integral scale
μ = kinematical viscosity
ρ = specific mass
ρ_s = specific mass of solid

REFERENCES

1. S. Mackrle, V. Mackrle, I. Tesařík and V. Mičan. *Reactor for Water Treatment by Sludge Blanket (in Czech)*. Czechoslovak Patent No. 88634 (1952).
2. S. Makrle, V. Mackrle, O. Dračka and L. Paseka, *Clarifier for Water Treatment by Coagulation and Sludge Blanket Filtration*. Czechoslovak Patent No.123929, Canad. Patent No. 769769 (1963).
3. S. Mackrle and V. Mackrle, *Arrangement for Biological Cleaning of Organically Polluted Liquid Composed of Building Units*. Czechoslovak Patent No. 159812, US Patent No. 3627136 (1970).
4. S. Mackrle and V. Mackrle, *Method of Biological Activating Cleaning with Fluid Filtration* Czechoslovak Patent No. 27574, US Patent No. 5032276 (1989).
5. S. Mackrle, V. Mackrle and O. Dračka, *Method of Separating Suspension, in Particular for Waste Water Treatment, and an Apparatus for Performing the Same*. Czech Patent No. 295871, PCT/CZ02/00027, US Patent No. 7,087,175 B2, US Patent No. 7,303,686 B2, European Patent No. 1390306, Russian Patent No. 2316482, China Patent No. ZL 02 8 10066.2, India Patent No. 221545 (2001).
6. S. P. Parker, (ed.) *Fluid Mechanics Source Book, Science Reference Series*, McGraw-Hill, New York (1988).
7. P. C. Carman, Fluid flow through granular bed. *Transactions of the Institute of Chemical Engineers (London)*, 15, 150–166 (1937).
8. V. Mackrle, *L'Étude du phénoméne d'adhérence. Colmatage dans le milieu poroux*. Théses, Faculté des Sciences de L'Université de Grenoble. Éditions de L'Académie des Sciences Tchecoslovaque, Section technique, Prague (1961) (1960).
9. J. O. Hinze, *Turbulence*, McGraw-Hill, New York (1959).
10. H. R. Kruyt, (ed.) *Colloid Science*, Vol. I., McGraw-Hill, New York (1959).
11. R. Shinnar, On the behavior of liquid dispersions in mixing vessels. *Journal of Fluid Mechanics*, 10, 259–275 (1961).
12. J. F. Davison and D. Harrison, (eds.), *Fluidization*, Academic, London (1971).
13. Metcalf and Eddy, Inc. *Wastewater Engineering—Treatment and Reuse*, 4th ed., McGraw-Hill, New York (2003).
14. L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas, (eds.), *Biological Treatment Processes*, Humana Press, Totowa, NJ, 818 pp. (2009).
15. L. K. Wang, N. K. Shammas and Y. T. Hung, (eds.), *Advanced Biological Treatment Processes*, Humana Press, Totowa, NJ (2009).
16. K. Imhoff and K. R. *Taschenbuch der Stadtenwässerung*, 28th ed., R. Oldenbourg, München (1993).
17. S. Mackrle, V. Mackrle and O. Dračka, *Small Capacity Waste Water Treatment Plant*. Czech Patent Applications PV 2001–3871 and PV 2002–3560, International Patent Application PCT/CZ02/00058 (Publ. No. WO 03/037805 A1) (2002).

11
Anaerobic Lagoons and Storage Ponds

Lawrence K. Wang, Yung-Tse Hung, and J. Paul Chen

CONTENTS

Abstract Anaerobic lagoons and storage ponds are frequently used for agricultural waste pre-treatment, and storage, respectively. This chapter introduces the anaerobic lagoon process, applications, limitations, performance, reliability, design considerations, energy consumption, capital cost, operation and maintenance costs, and process monitoring notes when the process is used for agricultural waste treatment. Practical design examples are presented in detail.

Key Words Anaerobic lagoons • storage ponds • agricultural waste • pre-treatment • sludge storage • applications • limitations • performance • reliability • design considerations • energy consumption • capital cost • operation and maintenance costs • process monitoring • design examples.

1. INTRODUCTION

In many situations, it is necessary to pretreat agricultural waste before final treatment. The purpose of pretreatment is to reduce pollution potential of the waste through biological, physical, and chemical processes. These types of components reduce nutrients, destroy pathogens,

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_11

and reduce total solids (TS). Pretreatment also includes solids separation, drying, and dilution that prepare the waste for facilitating another function. By the nature, pretreatment facilities require a higher level of management that that of waste storage facilities.

Anaerobic lagoons are frequently used for pretreatment of agricultural wastes. The lagoon effluent can be treated by various biological, physical, and chemical processes. This chapter introduces anaerobic lagoons, their applications, limitations, performance, reliability, deign considerations, energy consumption, capital costs, operation, and maintenance costs and process monitoring notes. Also discussed in this chapter are the waste storage ponds. Sometimes operators want to use lagoon effluent as flush water. To polish and store water for this purpose, waste storage ponds can be constructed in series with the anaerobic lagoon. The capacity of the waste storage pond should be sized for desired storage volume. The applications and design procedure are presented in detail.

2. PROCESS DESCRIPTION

Anaerobic lagoons are relatively deep (up to 20 ft) ponds with steep sidewalls in which anaerobic conditions are maintained by keeping loading so high that complete deoxygenation is prevalent. Although some oxygenation is possible in a shallow surface zone, once greases form an impervious surface layer, complete anaerobic condition develops. Treatment or stabilization results from thermophilic anaerobic digestion of organic wastes. The treatment processes is analogous to that occurring in single stage untreated anaerobic digestion of sludge in which acid forming bacteria break down organics. The resultant acids are then converted to carbon dioxide, methane, cells, and other end products.

In the typical anaerobic lagoon, raw wastewater enters near the bottom of the pond (often at the center) and mixed with the active microbial mass in the sludge blanket, which is usually about 6 ft deep. The discharge is located near one of the sides of the pond, submerges below the liquid surface. Excess undigested grease floats to the top, forming a heat retaining and relatively air tight over. Wastewater flow equalization and heating are generally not practiced. Excess sludge is washed out with the effluent. Recirculation of waste sludge is not required. Anaerobic lagoons are capable of providing treatment of high strength wastewaters and are resistant to shock loads. Figure 11.1 illustrates a typical anaerobic lagoon.

Anaerobic lagoons are customarily contained within earthen dikes. Depending on the soil characteristics, lining with various impervious material such as rubber, plastic, or clay may be necessary. Pond geometry may vary, but surface area to volume ratios is minimized to enhance heat retention (1–10).

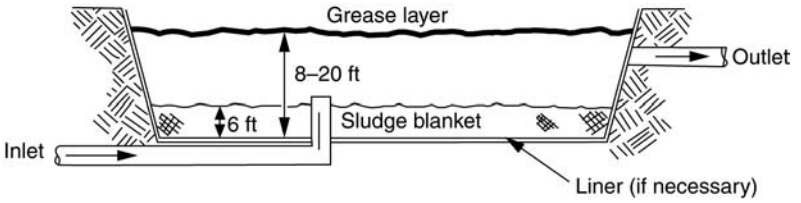


Fig. 11.1. Cross-section of a typical anaerobic lagoon.

3. APPLICATIONS AND LIMITATIONS

Although anaerobic biological processes are common for sludge digestion, anaerobic lagoons for wastewater treatment have found only limited applications. The anaerobic lagoon processes are well demonstrated for stabilization of highly concentrated organic wastes, such as animal wastes.

Anaerobes lagoons are currently accepted in the United States for the treatment of various animal wastes. Anaerobic treatment of animal waste helps to protect water quality by reducing much of the organic concentration (BOD, COD) of the waste. Anaerobic lagoons also reduce the nitrogen content of the waste through ammonia volatilization and effectively reduce animal waste odors if the lagoon is managed properly.

Anaerobic lagoons are also effective as treatment units prior to aerobic treatment of high strength wastes. Typically the anaerobic lagoons are used in series with aerobic or facultative lagoons. Anaerobic lagoon process may generate odors. It requires relatively large land area. For efficient operation, water temperature above 75°F should be maintained.

4. EXPECTED PROCESS PERFORMANCE AND RELIABILITY

BOD_5 removals of 50% to 70% are achievable depending on loading and temperature conditions. Total suspended solid (TSS) concentrations may increases, especially if the influent BOD_5 is primarily dissolved. Generally it doesn't produce as effluent suitable for direct discharge to receiving waters.

The process is generally resistant to upsets. It is highly reliable if pH in the relatively narrow optimum range is maintained. Anaerobic lagoons may create odors. The lagoons have relatively high land requirements. There is potential for changing wastewater into groundwater unless lagoon is lined.

In anaerobic lagoons, excess sludge is usually washed out in the effluent. Since anaerobic lagoons are often used for preliminary treatment recirculation or removal of sludge not generally required.

5. PROCESS DESIGN

5.1. Minimum Treatment Volume

The maximum operating level of an anaerobic lagoon is a volume requirement plus a depth requirement. The volume requirement is the sum of the following volumes:

1. Minimum required volume, ft^3 (MTV)
2. Manure volume and wastewater volume, ft^3 (WSV)
3. Sludge volume, ft^3 (SV)

Polluted runoff from a watershed must not be included in an anaerobic lagoon unless a conservative estimate of the volatile solid loading can be made. Runoff from a watershed such as a feedlot, is not included in an anaerobic lagoon because loading would only result during storm events and because the magnitude of the loading would be difficult, if not impossible, to estimate. As a result, the lagoon would be shocked with an overload of volatile solids.

An automatic outflow device, pipe, or spill way must be placed at a height above the maximum operating level to accommodate the following depths:

1. Normal precipitation less evaporation on lagoon surface, ft
2. The 25 year, 24-h storm precipitation on lagoon surface, ft

These depths added to the depth of the volume requirement of the lagoon determines the level of the outflow device, pipe, or spillway. The depth of head required to operate the outflow plus a minimum of 1 ft of freeboard is provided above the outflow and establishes the top of the embankment. Should state regulation preclude the use of an outflow device, pipe, or spillway or if for some other reason the lagoon will not have these, storing the 25 year, 24-h storm precipitation on the anaerobic lagoon surface (a second time) replaces the head requirement.

The combinations of these volumes and depths are illustrated in Fig. 11.2. The terms and derivation are explained in the following paragraphs.

Anaerobic lagoons are designed on the basis of volatile solids loading rate (VSLR) per 1000 ft^3. Volatile solids represent the amount of solids material in the wastes that will decompose as opposed to the mineral (inert) fraction. The rate of solids decomposition in anaerobic lagoons is a function of temperature; therefore, the acceptable VSLR varies from one location to another. Figure 11.3 indicates the maximum VSLRs for the United States. If odors need to be minimized, VSLR should be reduced by 25% to 50%.

The minimum treatment volume (MTV) represents the volume needed to maintain sustainable biological activity. The minimum treatment volume for VS can be determined using Eq. (1).

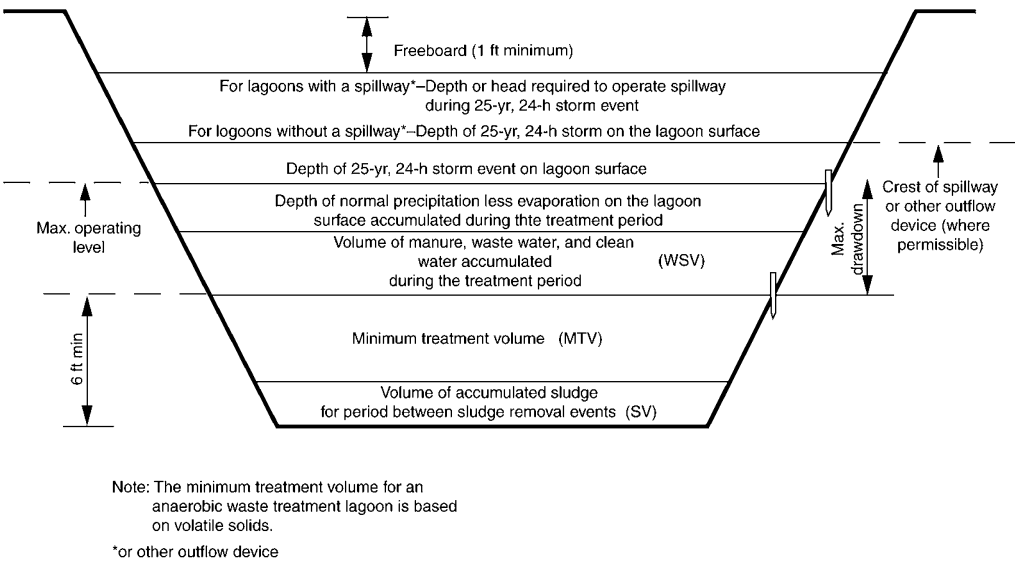


Fig. 11.2. Illustration of volumes and depth requirements for anaerobic lagoons.

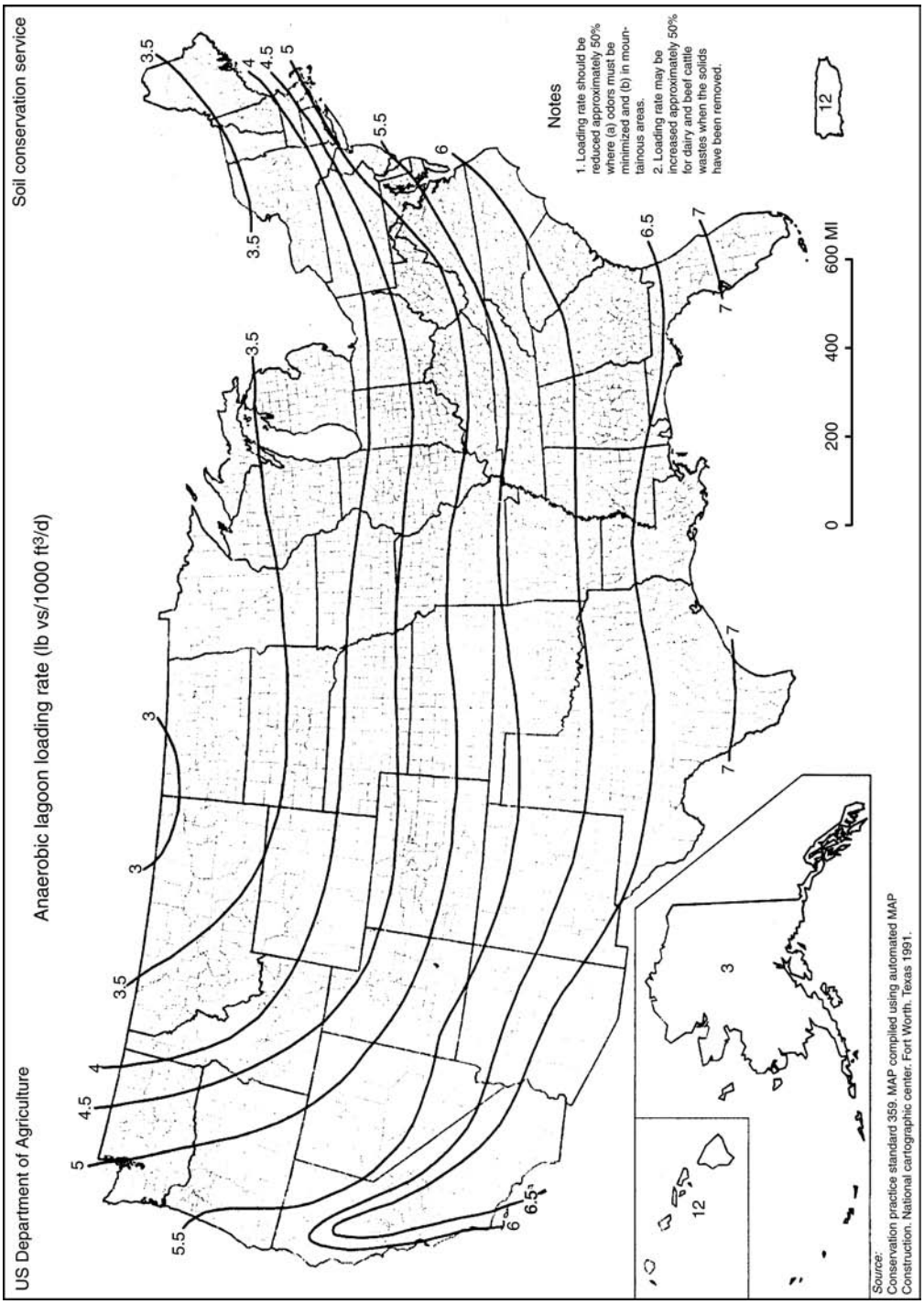


Fig. 11.3. Anaerobic lagoons volatile solids loading rates in USA.

$$\text{MTV} = \frac{\text{TVS}}{\text{VSLR}} \tag{1}$$

where

MTV = minimum treatment volume, ft^3
TVS = total daily volatile solids loading (from all sources), lb/day
VSLR = volatile solids loading rate, lb/1000 ft^3/day (this can be obtained from Fig. 11.3)

5.2. *Waste Volume for Treatment Period*

Daily volatile solids production for various wastes can be determined using waste volume (WV) for treatment period. If feed spillage exceeds 5%, VSP should be increased by 4% for each additional 1% spillage.

Waste volume (WV) should reflect the actual volume of manure, waste water, flash water that will not be recycled, and clean dilution water added to the lagoon during the treatment period. The treatment period is either the detention time required to obtain the desired reduction of pollution potential of the waste or the time between land application events, whichever is longer. State regulations may govern the minimum detention time. Generally, the maximum time between land application events determines the treatment period because this time generally exceeds the detention time required.

$$\text{WV} = \text{TVM} + \text{TWW} + \text{CW} \tag{2}$$

where

WV = waste volume for treatment period, ft^3
TVM = total volume of manure for treatment period, ft^3
TWW = total volume of wastewater for treatment period, ft^3
CW = clean water added during treatment period, ft^3

5.3. *Sludge Volume*

As the manure is decomposed in the anaerobic lagoon only part of the total solids (TS) is reduced. Some of the TS are mineral material that will not decompose, and some of the VS require a long time to decompose. These materials, referred to as sludge, gradually accumulate in the lagoon. To maintain the minimum treatment volume (MTV), the volume of sludge accumulation over the period of time between sludge removals must be considered. Anaerobic lagoons are commonly designed for 15 to 20 years sludge accumulation period. The sludge volume (SV) can be determined using Eq. (3).

$$\text{SV} = 365 \times \text{AU} \times \text{TS} \times \text{SAR} \times T \tag{3}$$

where

SV = sludge volume, ft^3
AU = number of 1000-lb animal units
T = sludge accumulation time, years
TS = total solids production per animal unit per day, lb/AU/day
SAR = sludge accumulation ratio, ft^3/lb TS

5.4. Lagoon Volume Requirement

Total solids values can be obtained from the site investigations. Sludge accumulation ratios should be taken from Table 11.1. An SAR is not available for beef, but it can be assumed as similar to that for dairy cattle.

The lagoon volume requirements are for accommodation of the minimum pretreatment volume, the sludge volume, and the waste volume for the treatment period. This is expressed in Eq. (4).

$$LV = MTV + SV + WV \tag{4}$$

where

LV = lagoon volume requirement, ft^3
MTV = minimum treatment volume, ft^3, see Eq. (1)
SV = sludge volume accumulation for period between sludge removal events, ft^3, see Eq. (3)
WV = waste volume for the treatment period, ft^3, see Eq. (2)

In addition to the anaerobic lagoon volume requirement (LV), a provision must be made for depths to accommodate the normal precipitation less evaporation on the anaerobic lagoon surface; the 25-year, 24-h storm precipitation; the depth required to operate the emergency outflow; and freeboard. Normal precipitation on the lagoon surface is based on the critical treatment period that produces the maximum depth. This depth can be offset to some degree by evaporation losses on the lagoon surface. The offset varies, according to the climate of the region, from a partial amount of the precipitation to an amount in excess of the precipitation. Precipitation and evaporation can be determined from local climate data. Figure 11.4 shows the average evaporation data in the United States.

The minimum acceptable depth for anaerobic lagoon is 6 ft, but in cold climate at least 10 ft is recommended to assure proper operation and odor control. The design height of an embankment for an anaerobic lagoon should be increased by the amount needed to ensure that the design elevation is maintained after settlement. This increase should not be less than 5% of the design fill height. The minimum top width of the lagoon should be as shown in Table 11.2.

The combined side slopes of the settle embankment should not be less than 5 to 1 (horizontal to vertical). The inside slopes can vary from 1 to 1 for excavated slopes to 3 to 1 or flatter where embankments are used. Construction technique and soil type must also be considered.

Table 11.1
Sludge accumulation ratios

Animal type	Sludge accumulation ratio
Poultry	
Layers	0.0295
Pullets	0.0455
Swine	0.0185
Dairy cattle	0.0729

Source: USDA.

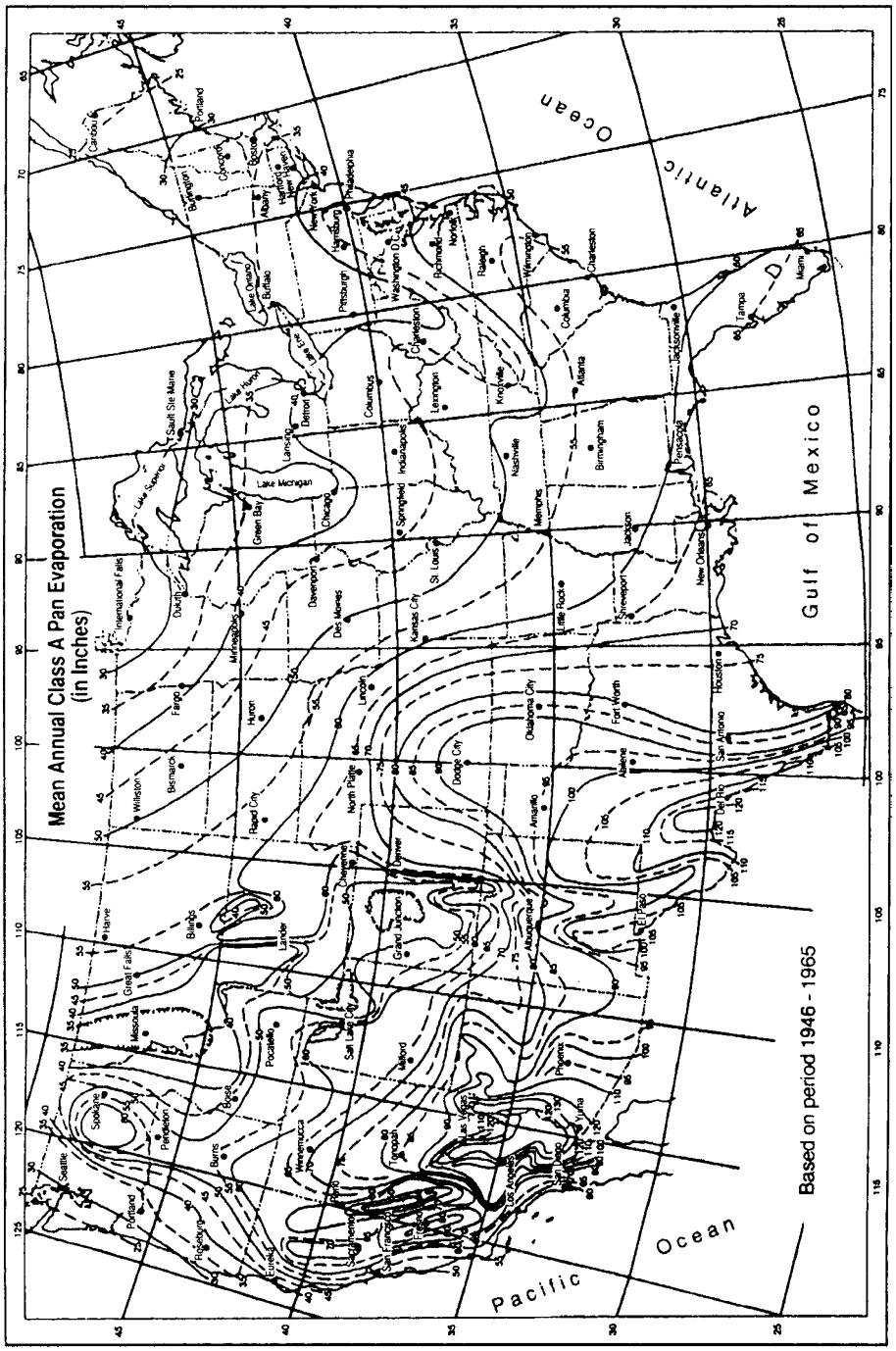


Fig. 11.4. Average evaporation data in the United States.

Table 11.2
Minimum top width for lagoon embankments

Maximum length of embankment (ft)	Top width (ft)
10 or less	6
11–14	8
15–19	10
20–24	12
25–34	14
34 or more	15

Source: USDA.

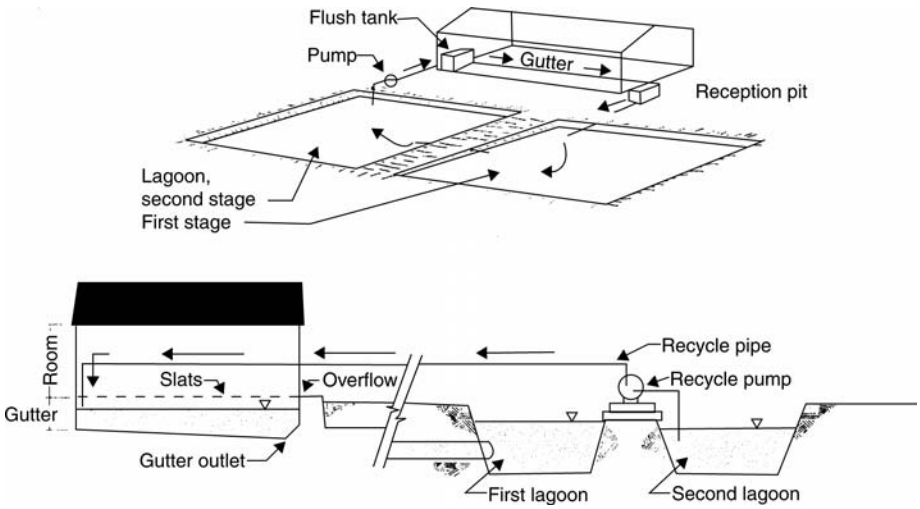


Fig. 11.5. Two anaerobic lagoons with recycle system.

In some situation a steep slope may be used below the design liquid level, while a flatter slope is used above the liquid level to facilitate maintenance and bank stabilization. The minimum elevation of the top of the settled embankment should be 1 ft above the maximum design water surface of the lagoon.

5.5. Anaerobic Lagoon Design Criteria

Important criteria for designing an anaerobic lagoon system are summarized in below. Figure 11.5 shows a two-lagoon system.

1. Operation: parallel or series.
2. Detention time: 20 to 180 days.
3. Depth: 8 to 20 ft.
4. pH: 6.8 to 7.2.
5. Water temperature range: 35°F to 120°F.
6. Optimum waste temperature: 86°F.
7. Organic loading 200 to 2200 lb BOD_5/acre/day.

8. Nutrient requirement: Nutrient as needed to make up deficiencies in raw wastewater. No other chemical required.
9. Leakage prevention: A lagoon should be constructed to avoid leakage and potential in ground water pollution.
10. Overtopping prevention: If overtopping can cause embankment failure, an emergency spillway or over flow pipe should be provided. A lagoon can have an over flow to maintain a constant liquid level if the overflow liquid is stored in waste storage pond or otherwise properly managed.
11. Inlet anti-freezing protection.
12. Sludge removal: Sludge removal is an important consideration in the design. This can be accomplished by agitating the lagoon and pumping out the mixed sludge or by using a drag- line for removal floating or settled sludge.

5.6. Data Gathering and Compilation for Design

Anaerobic lagoons can be used for treatment of both animal wastes and wastewater. The major application of anaerobic lagoons however is for animal waste treatment.

In case an agricultural waste treatment system is to be developed for an animal farm, the following information should be gathered:

1. Type of animal
2. Design population of animal
3. Average weight of each animal (lb)
4. 25-year 24-h storm for the local area (in.)
5. Net precipitation = precipitation - evaporation (in.)
6. Time interval between lagoon pumping = treatment period (days)
7. Time interval between sludge removal (years)
8. Daily volume of daily manure production (ft^3/AU/day)
9. Daily wastewater volume per animal unit (ft^3/AU/day)
10. Clean water added during treatment period (ft^3)
11. Daily manure to all solids production (lbs/AU/day)
12. Percent volatile content in the total solid manure (%)
13. Lagoon volatile solids loading rate (lb VS/1000 ft^3)
14. Sludge accumulation ratio (ft^3/lb TS)
15. Sludge accumulation period (years)
16. Anaerobic lagoon's side slope ratio (horizontal to vertical ratio)

6. ENERGY CONSUMPTION AND COSTS OF ANAEROBIC LAGOONS

Anaerobic lagoons are operated by gravity glow and therefore have no energy requirement other than any pumping that may be necessary to lift the influent wastewater into the lagoons.

Table 11.3 shows the construction costs (January 2002 dollars; ENR CC index = 6390.21) and Table 11.4 shows the operation and maintenance cost under the following engineering assumptions:

1. January 2002 US dollars
2. Service life: 50 years
3. Average detention time = 35 days

Table 11.3
Construction cost of anaerobic lagoons

Wastewater flow (MGD)	Construction cost US10^6(January 2002)
0.1	0.1246
0.5	0.4005
1.0	0.7120
5.0	2.4475
10.0	3.8938
50.0	13.3500
100.0	22.2500

Source: US EPA.

Table 11.4
Operation and maintenance cost of anaerobic lagoons

Waste water flow (MGD)	Annual O & M Costs US$ 10^6(January 2002)
0.1	0.0040
0.5	0.0100
1.0	0.0156
5.0	0.0445
10.0	0.0690
50.0	0.2069
100.0	0.3115

Source: US EPA.

4. Depth = 10 ft
5. BOD_5 loading = 466 lb/acre/day
6. Constriction cost includes excavating, grading, and other earthwork and service roads.
7. Costs don't include land and pumping
8. Operation and maintenance cost consist of labor and material
9. Waste water characteristics: influent BOD_5 = 600 mg/L; effluent BOD_5 = 240 mg/L

To adjust costs for other BOD_5 loading and/or detention times, enter the tabulated data at effective flow

$$Q = Q_{\text{design}} \times \frac{(466\,\text{lb/acre/day})(\text{new detention time})}{(\text{new design loading})(35\,\text{days})} \tag{5}$$

It should be noted that the above data of construction cost and O & M cost have been complied by the US Environmental Protection Agency for wastewater treatment using anaerobic lagoons. Whether or not the same data can be applied to animal waste treatment remains unknown.

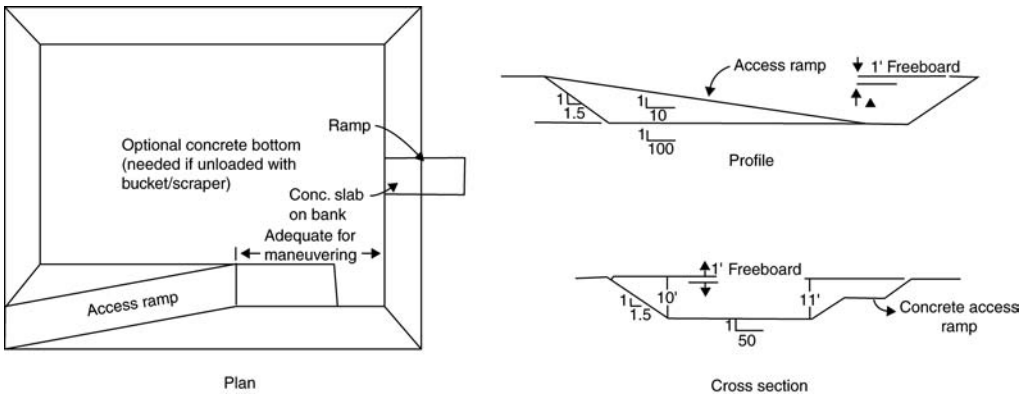


Fig. 11.6. Layout of waste storage pond.

7. WASTE STORAGE PONDS

7.1. Process Description

Sometimes operators want to use lagoon effluent as flush water. To polish and store water for this purpose, waste storage ponds can be constructed in series with the anaerobic lagoon. Storage ponds are earthen basins designed to store wastewater, sludge, and manure (Fig. 11.6). They generally are rectangular, but may be circular or any other shape that is practical for operation and maintenance. The capacity of the waste storage pond should be sized for the desired storage volume. A minimum capacity of the waste storage ponds is the volume for rainfall (RFV), runoff (ROF), and emergency storm storage (ESV). By limiting the depth to less than 6 ft, the pond will function more nearly likes an aerobic lagoon. Odors and the level of ammonia, ammonium, and nitrate will be more effectively reduced.

Earthen storage is frequently the least expensive type of storage of sludge and manure, however, certain restrictions, such as limited space availability, high precipitation, water table, permeable soils, or shallow bedrock, can limit the types of storage considered.

7.2. Process Design

Liquid waste storage ponds and structures should be sized to hold all of the manure, bedding, and wastewater from milkhouse, flushing, and contaminated runoff that can be expected during the storage period. Equation (6) can be used to compute the waste storage volume:

$$\mathrm{WSV} = \mathrm{TVM} + \mathrm{TWW} + \mathrm{TBV} + \mathrm{CW} + \mathrm{ROV} + \mathrm{VSA} \tag{6}$$

where

WSV = waste storage volume for storage periods, ft^3
TVM = total volume of manure for storage period, ft^3
TWW = total wastewater volume for storage period, ft^3

TBV = total bedding volume for storage period, ft^3
CW = clean water added during storage period, ft^3
ROV = runoff volume, ft^3
VSA = solids accumulation volume, ft^3

Figure 11.7 shows the cross section of a waste storage pond without a watershed; while Fig. 11.8 shows the cross section of a waste storage pond with a watershed. Various parameters such as ROV, TVM, CW, TWW, and VSA are clearly illustrated.

In addition to the waste storage volume, waste storage facilities must, if uncovered, provide a depth to accommodate precipitation less evaporation on the storage surface during the most critical storage period. The most critical storage period is generally the consecutive months that represent the storage period that gives the greatest depth of the precipitation less evaporation.

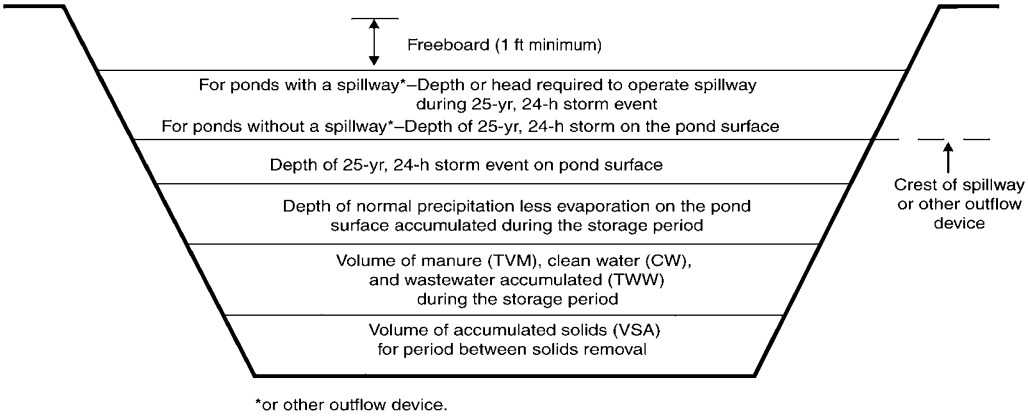


Fig. 11.7. Cross section of a waste storage pond without a watershed.

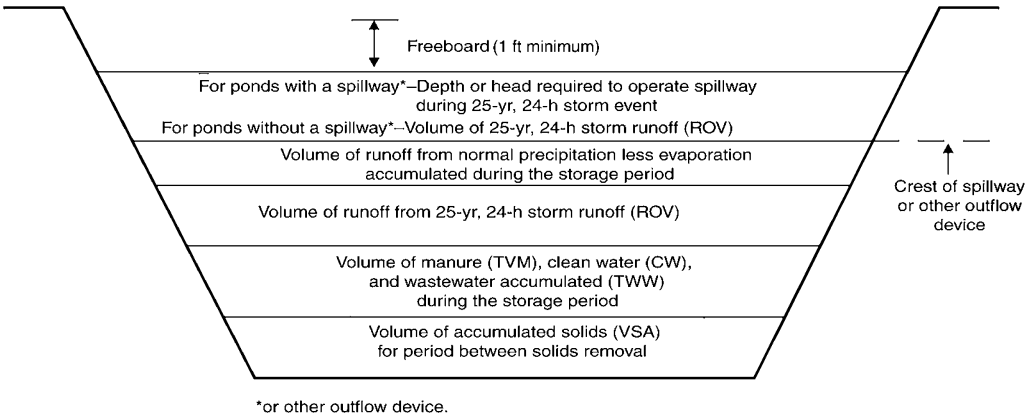


Fig. 11.8. Cross section of a waste storage pond with a watershed.

Frequently, waste storage ponds are designed to include outside runoff from watershed. For these, the runoff volume of the 25-year, 24-h storm must be included in the storage volume. if the pond does not have a spillway or other outflow device, the runoff volume of 25-year, 24-h storm must be included a second time.

Accordingly the total depth of a waste storage pond can be estimated as summation of the following:

1. Pond depth calculated based on minimum storage volume (WSV)
2. Added depth due to "precipitation less evaporation" for the storage period
3. Added depth due to 25-year, 24-h storm (only for the ponds without a drainage area)
4. Added depth required to operate emergency outflow
5. Added depth for freeboard (1 ft minimum)

8. DESIGN AND APPLICATION EXAMPLES

8.1. Example 1

For lagoon sizing and design, how can the volume of a rectangular lagoon be calculated?

Solution:

The rectangular lagoon volume can be calculated by the following Eq. (7)

$$V = \frac{\left(4 \times Z^2 \times d^3\right)}{3} + \left(Z \times \mathrm{BL} \times d^2\right) + \left(Z \times \mathrm{BW} \times d^2\right) + (\mathrm{BW} \times \mathrm{BL} \times d) \qquad (7)$$

where

V = lagoon volume, ft^3
Z = side slope ratio (horizontal to vertical)
d = lagoon depth, ft
BW = lagoon bottom width
BL = lagoon bottom length

8.2. Example 2

How can the volume of a circular lagoon be calculated?

Solution:

The circular lagoon volume can be determined by Eq. (8):

$$V = \left(1.05 \times Z^2 \times d^3\right) + \left(1.57 \times W \times Z \times d^2\right) + \left(0.79 \times W^2 \times d\right) \qquad (8)$$

where

V = lagoon volume, ft^3
Z = side slope ratio
d = lagoon depth, ft
W = lagoon bottom diameter, ft

8.3. Example 3

Develop a step-by-step design procedure for designing an anaerobic lagoon system to treat the manures from an agricultural farm.

Solution:

Step1. Determine animal units

1a. Animal type.....
1b. Animal weight (W).....lb
1c. Number of animal (N).....
1d. Animal units (AU) = W (N)/1000 = ...

Step 2. Determine manure volume

2a. Daily volume of manure production per AU (DVM) = ...ft^3/AU/day
2b. Treatment period (D) = ...day
2c. Total volume of manure production for animal type and treatment period VMD = AU × DVM × D = ...ft^3
2d. Total manure production for treatment period (TVM) = ...ft^3

Step 3. Determine wastewater volume

3a. Daily wastewater volume per AU (DWW) = ...ft^3/AU/day
3b. Total wastewater volume for animal description and treatment period (WWD) = DWW × AU × D = ...ft^3
3c. total wastewater volume for treatment period (TWW) = ... ft^3

Step 4. Determine clean water volume

4a. Clean water added during treatment period (CW) = ... ft^3

Step 5. Determine waste volume

5a. Waste volume for treatment period (WV) = TVM + TWW + CW

= ..... + ... + ...

= .....ft^3

Step 6. Determine the manure total solids production

6a. Daily manure total solids production (MTS) = ... lb/AU/day
6b. Daily manure total solids production for animal type (MTSD)

= MTS × AU = ...lb/day

6c. Total manure total solids production (TMTS) = ...lb/day

Step 7. Determine manure volatile solids production

7a. Daily manure volatile solids production per AU (MVS) = ...lb/AU/day
7b. Daily manure volatile solids production for animal type per day (MVSD) = AU × MVS = ...lb/day

7c. Total manure volatile solids production (TMVS) = ...lb/day

Step 8. Determine wastewater volatile solids production

8a. Daily wastewater volatile solids production (DWVS) = ...lb/1000 gal
8b. Total wastewater volatile solids production for animal type (WVSD) = DWVS × DWW × 7.48/(D × 1000) = ...lb/day
8c. Total wastewater volatile solids production (TWVS) = ... lb/day

Step 9. Determine total volatile solids (manure and wastewater) production

9a. Total daily volatile solids production (TVS) = TMVS + TWVS = ... + ... = ...lb/day

Step 10. Determine minimum treatment volume

10a. Select lagoon VS loading rate (VSLR) = ... lb VS/1000 ft^3/d
10b. Minimum treatment volume (MTV) = TVS × 1000/ VSLR
= . . . . . × . . . . . /(. . . . .) = ... ft^3

Step 11. Determine sludge volume requirement

11a. Sludge accumulation ratio (SAR) = ... ft^3/lb TS
11b. Sludge accumulation period (T) = ...year
11c. Sludge volume requirement (SV) = 365 × TMTS × T × SAR
= 365 × . . . . . × . . . × . . . . . = ... ft^3

Step 12. Determine minimum lagoon volume requirements

12a. Minimum lagoon volume requirement (MLVR) = MTV + SV + WV
= . . . . . + ... + ... = ... ft^3

Step 13. Determine lagoon size

13a. Side slope ratio (Z) = ...
13b. Lagoon volume

$$V = \left[\left(4 \times Z^2 \times d^3\right)/3\right] + \left(Z \times \mathrm{BL} \times d^2\right) + \left(Z \times \mathrm{BW} \times d^2\right) + (\mathrm{BW} \times \mathrm{BL} \times d)$$

13c. Lagoon volume (V) must be equal to or greater than MLVR = ...ft^3
13d. Determine the closest lagoon volume

Trial 1. BW = ... ft^3; BL = ... ft; d = ... ft;
V = ... ft^3
Trial 2. BW = ...ft^3; BL = ... ft; d = ...ft**;**
V = ... ft^3
Trail 3. BW = ... ft^3; BL = ...ft; d = ...ft;
V = ... ft^3
Select V = ...ft^3 ≈ MLVR

Step 14. Depth adjustment

14a. Depth (d) = ... ft
14b. Add depth of precipitation less evaporation on lagoon surface for the treatment period = ...ft
14c. Add depth of 25-year, 24-h storm = ...ft
14d. Add depth required to operate emergency outflow (note: if lagoon design doesn't include a spillway or other automatic outflow device, use depth of 25-year, 24-h storm precipitation = ...ft
14e. Add for freeboard (1 ft minimum) = ...ft
14f. Final depth = ...ft; use ..... ft

Step 15. Compute total volume of rectangular anaerobic lagoon using final depth

$$V = \left[\left(4 \times Z^2 \times d^3\right)/3\right] + \left(Z \times \mathrm{BL} \times d^2\right) + \left(Z \times \mathrm{BW} \times d^2\right) + (\mathrm{BW} \times \mathrm{BL} \times d)$$
$$= \ldots\ldots \mathrm{ft}^3$$

8.4. Example 4

An animal farm has formally requested assistance in developing an agricultural waste treatment system using an anaerobic lagoon. Assuming you are an environmental engineer, design an anaerobic lagoon system for the animal firm based on the flowing given information:

1. Type of animal = pigs
2. Design population of animal = 6000
3. Average weight of animal = 150 lb
4. 25-year, 24-h storm for the local area = 6 in
5. Net precipitation = precipitation - evaporation = 2 in
6. Time interval between lagoon pumping = treatment period = 180 day
7. Time interval between sludge removal = 5 year
8. Daily volume of daily manure production = 1 ft^3/AU/day
9. Daily wastewater volume per animal unit = 0 ft^3/AU/day
10. Clean water added during treatment period = 0 ft^3
11. Daily manure solids production = 6.34 lb/AU/day
12. Percent volatile content in the total solids of manure = 85.17%
13. Lagoon volatile solids loading rate = 6 lb VS/1000 ft^3
14. Sludge accumulation ratio = 0.0485 ft^3/lb TS
15. Sludge accumulation period = 5 year
16. Anaerobic lagoon's side slope ratio (horizontal to vertical ratio) = 2

Solution:

Since 85.11% of the total solids in the given information item (R) is volatile, the daily manure volatile solids production per AU, or MVS, is estimated to be 5.4 lb/AU/day.

The step-by-step design procedures are used for the detailed design as follows:

Step 1. Determine animal units

1a. Animal type pigs
1b. Animal weight (W) 150 lb

1c. Number of animals (N) 6000
1d. Animal units (AU) = W (N)/1000 = 900

Step 2. Determine manure volume

2a. Daily volume of manure production per AU (DVM) = 1.0 ft^3/AU/day
2b. Treatment period (D) = 180 days
2c. Total volume of manure production for animal type and treatment period

$$VMD = AU \times DVM \times D = 162{,}000\ ft^3$$

2d. Total manure production for treatment period (TVM) = 162,000 ft^3

Step 3. Determine wastewater volumes

3a. Daily wastewater volume per AU (DWW) = 0 ft^3/AU/day
3b. Total wastewater volume for animal description and treatment period (WWD) = DWW × AU × D = 0 ft^3
3c. Total wastewater volume treatment period (TWW) = 0 ft^3

Step 4. Determine clean water volume

4a. Clean water added during treatment period (CW) = 0 ft^3

Step 5. Determine waste volume

5a. Waste volume for treatment period (WV) = TVM + TWW + CW = 162,000 + 0 + 0 = 162,000 ft^3

Step 6. Determine the manure total solids production

6a. Daily manure total solids production (MTS) = 6.34 lb/AU/day
6b. Daily manure total solids production for animal type (MSTD) = MTS × AU = 5706 lb/day
6c. Total manure total solids production (TMTS) = 5706 lb/day

Step 7. Determine manure volatile solids production

7a. daily manure volatile solids production per AU (MVS) = 5.4 lb/AU/day
7b. Daily manure volatile solids production for animal type per day (MVSD) = AU × MVS = 4860 lb/day
7c. Daily manure volatile solids production = (TMVS) = 4860 lb/day

Step 8. determine waste water volatile solids production

8a. Daily waste water volatile solids production (DWVS) = 0 lb/1000 gal
8b. Total wastewater volatile solids production for animal type (WVSD) = DWVS × DWW × 7.48/(D × 1000) = 0 lb/day
8c. Total wastewater volatile solids production (TWVS) = 0 lb/day

Step 9. Determine total volatile solids (manure and wastewater) production

9a. Total daily volatile solids production (TVS) = TMVS + TWVS = 4860 + 0 = 4860 lb/day

Step 10. Determine minimum treatment volume

10a. Select lagoon VS loading rate (VSLR) = 6 lb VS/1000 ft^3/d
10b. Minimum treatment volume (MTV) = TVS × 1000/VSLR = 4860 × 1000/6 = 810,000 ft^3

Step 11. Determine sludge volume requirement

11a. Sludge accumulation ratio (SAR) = 0.0485 ft^3/lb TS
11b. Sludge accumulation period (T) = 5 year
11c. Sludge volume requirement (SV) = 365 × TMTS × T × SAR = 365 × 5706 × 5 × 0.485 = 505,052 ft^3

Step 12. Determine minimum lagoon volume requirement

12a. Minimum lagoon volume requirement (MLVR) = MTV + SV + WV = 810,000 + 505,052 + 162,000 = 1,477,052 ft^3

Step 13. Determine lagoon size

13a. Side slope ratio (Z) = 2
13b. Lagoon volume (V)= $[(4 \times Z^2 \times d^3)/3] + (Z \times BL \times d^2) + (Z \times BW \times d^2) + (BW \times BL \times d)$
13c. Lagoon volume (V) must be equal to or greater than MLVR = 1,477,052 ft^3
13d. Determine the closest lagoon volume
Trail 1. BW = 150 ft; BL = 1000 ft; d = 8 ft; V = 1,344,931 ft^3
Trail 2. BW = 150 ft; BL = 1200 ft; d = 8 ft; V = 1,615,531 ft^3
Trial 3. BW = 150 ft; BL = 1100 ft; d = 8 ft; V = 1,482,731 ft^3
Select V = 1,482,731 $ft^3 \approx$ MLVR

Step 14. Depth adjustment

14a. Depth (d) = 8 ft
14b. Add depth of precipitation less evaporation on lagoon surface for the treatment period = 0.6 ft
14c. Add depth of 25-year, 24-h storm = 0.5 ft
14d. Add depth required to operate emergency outflow (note: if lagoon deign does not include a spillway or other automatic outflow device, use depth of 25-year, 24-h storm precipitation) = 0.3 ft
14e. Add for freeboard (1 ft minimum) = 1.0 ft
14f. Final depth = 10.4 ft; use 10.5 ft

Step 15. Compute total volume of rectangular anaerobic lagoon using final depth

$$(V) = [(4 \times Z^2 \times d^3)/3] + (Z \times BL \times d^2) + (Z \times BW \times d^2) + (BW \times BL \times d)$$
$$= 2{,}014{,}300\ ft^3$$

8.5. Example 5

An animal farm has 500 milkers animals each weighing 1400 lb, 150 dry animal each weighing 1400 lb and 150 Heifers animals each weighing 1000 lb. The daily volume of manure production rates (DVM) are:

(a) Milkers = 1.3 ft^3/AU/day

(b) Dry = 1.1 ft^3/AU/day
(c) Heifers = 1.3 ft^3/AU/day

Determine (a) the total animal units (AU), and (b) the total manure production for intended storage period of 180 days.

Solution:

1. *Determination of total animal units*

$$\begin{aligned} AU &= W_1N_1/1000 + W_2N_2/1000 + W_3N_3/1000 \\ &= (1400 \times 500)/1000 + (1400 \times 150)/1000 + (1000 \times 150)/1000 \\ &= 700 + 210 + 150 = AU_1 + AU_2 + AU_3 \end{aligned}$$

2. *Determination of total manure production (TVM)*

$$\begin{aligned} TVM &= AU_1 \times DVM_1 \times D + AU_2 \times DVM_2 \times D + AU_3 \times DVM_3 \times D \\ &= 700 \times 1.3 \times 180 + 210 \times 1.1 \times 180 + 150 \times 1.3 \times 180 = 163800 + 41580 + 35100 \\ &= 240{,}480\ ft^3 \end{aligned}$$

8.6. Example 6

Determine the waste storage volume for designing a waste storage pond for the same animal farm described in Example 5. Assuming the following is known:

(a) Daily wastewater volume per AU for Milkers = 0.6 ft^3 /AU/day
(b) Daily waste water volume per AU for Dry = 0
(c) Daily waste water volume per AU for Heifers = 0

The clean water volume (CW), runoff volume (ROV) and solids accumulation volume (VSA) are all zero. Total manure production (TVM) is 240,480 ft^3 from Example 5.

Solution:

Total wastewater volume for the storage period (TWW) can be determined as follows:

$$\begin{aligned} TWW &= WWD_1 + WWD_2 + WWD_3 \\ &= DWW_1 \times AU_1 \times D + DWW_2 \times AU_2 \times D + DWW_3 \times AU_3 \times D \\ &= 0.6 \times 700 \times 180 + 0 \times 210 \times 180 + 0 \times 150 \times 180 = 75,600\ ft^3 \end{aligned}$$

Waste storage volume (WSV) can then be calculated using Eq. (6)

$$\begin{aligned} WSV &= TVM + TWW + TBV + CW + ROV + VSA \\ &= 240{,}480 + 75{,}600 + 0 + 0 + 0 + 0 = 316{,}080\ ft^3 \end{aligned}$$

8.7. Example 7

Based on the technical information from Examples 5 and 6, and design a rectangular waste storage pond, assuming:

(a) Waste storage volume (WSV) = 316,080 ft^3
(b) Side slope ratio (Z) = 3
(c) Depth of precipitation less evaporation for the storage period = 2.3 ft
(d) 25-year, 24-h storm (for ponds without a drainage area) = 0.3 ft
(e) Depth requirement to operate emergency outflow = 0.3 ft
(f) Freeboard requirement = 1 ft minimum

Solution:

Equation (7) in Example 1 is used for calculating pond volume (V).

Trial 1. BW = 100 ft; BL = 500 ft; d = 6 ft; V = 367,392 ft^3
Trial 2. BW = 100 ft; BL = 400 ft; d = 6 ft; V = 296,592 ft^3
Trial 3. BW = 100 ft; BL = 425 ft; d = 6 ft; V = 314,292 ft^3
Trial 4. BW = 100 ft; BL = 428 ft; d = 6 ft; V = 320,580 ft^3

Very close to WSV of 316,080 ft^3.

Finally the waste storage pond's depth must be adjusted in order to determine the final depth:

$$\text{Final depth} = 6.1 + 2.3 + 0.3 + 0.3 + 1 = 10\,\text{ft}$$

NOMENCLATURE

AU = number of 1000-lb animal units
BL = lagoon bottom length
BW = lagoon bottom width
CW = clean water added during treatment period, ft^3
d = lagoon depth, ft
LV = lagoon volume requirement, ft^3
MTV = minimum treatment volume, ft^3, see Eq. (1)
N = number of animals
ROV = runoff volume, ft^3
SAR = sludge accumulation ratio (ft^3/lb TS)
SV = sludge volume (ft^3)
T = sludge accumulation time (year)
TBV = total bedding volume for storage period, ft^3
TS = total solids production per animal unit per day (lb/AU/day)
TVM = total volume of manure for treatment period, ft^3
TVS = total daily volatile solids loading (from all sources), lb/day
TWW = total wastewater volume, ft^3
V = lagoon volume, ft^3
VSA = solids accumulation volume, ft^3
VSLR = volatile solids loading rate, lb/1000 ft^3/day
W = lagoon bottom diameter, ft
W = average weight of each animal, lb

WSV = waste storage volume for storage periods, ft^3
WV = waste volume for the treatment period, ft^3
Z = side slope ratio (horizontal to vertical)

REFERENCES

1. US EPA, *Innovative and Alternative Technology Assessment*. 430/9-78-009. US Environmental Protection Agency, Washington, DC (1980).
2. US EPA, *Municipal Wastewater Stabilization Ponds*. EPA-625/1-83-015. US Environmental Protection Agency, Washington, DC (1983).
3. C. P. C. Poon, L. K. Wang, and M. H. S. Wang, Waste Stabilization Ponds and Lagoons, Chapter 7. In: *Handbook of Environmental Engineering*, *Volume 3. Biological Treatment Processes*, L. K. Wang and N. C. Pereira (eds.), pp. 305–360, Humana Press, Totowa, NJ (1986).
4. D. H. Vanderholm, D. L. Day, A. J. Muehling, and Y. T. Hung, *Management of Livestock Wastes*. PB 2000-100255. US Department of Commerce, National Technical Information Service, Springfield, VA, June (2000).
5. M. Krofta and L. K. Wang, *Treatment of Household Wastes, Septage and Septic Tank Effluents*. PB2000-101750INZ. US Department of Commerce, National Technical Information Service, Springfield, VA, November (2000).
6. US EPA, *Control of Air Emission from Superfund Sites*. EPA-625/R-92-012. US Environmental Protection Agency, Washigton, DC, p 119 (1992).
7. NSFC, Ohio Livestock Manure and Waste Water Management Guide. WWBKMGIO. *Small Flows*, 4, 1, 46. National Small Flows Clearing House, West Virginia University. Morgantown, WV (2003).
8. Editor, 2002–2003. Buyer's guide: Lagoon equipment, *Water Environment and Federation*, 14, 93–94 (2002).
9. L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.) Biosolids Treatment Processes. Humana Press, Totowa, NJ. 820 p. (2007).
10. L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.) Biosolids Engineering and Management. Humana Press, Totowa, NJ. 800 p. (2008).

12
Vertical Shaft Digestion, Flotation, and Biofiltration

Lawrence K. Wang, Nazih K. Shammas, Jeffrey Guild, and David Pollock

CONTENTS

Abstract Three new biosolids treatment processes are introduced: vertical shaft digestion (VSD), vertical shaft floatation (VSF) thickening, and gas-phase biofiltration. The combination of these three major processes, and a few supplemental units (such as grit removal, dewatering and drying) provide complete biosolids treatment. The topics covered in this chapter include: biosolids treatment objectives, vertical shaft bioreactor, vertical shaft flotation, vertical shaft digestion, aerobic digestion, autothermal thermophillic aerobic digestion, anaerobic digestion, dewatering, air emission control by biofiltration, engineering design, and case histories.

Key Words Vertical shaft bioreactor • wastewater treatment • vertical shaft digestion • sludge treatment • flotation • sludge thickening • biofiltration • air emission control • design • case histories.

1. INTRODUCTION

1.1. Biosolids Treatment

Solids processing represents about 40% of the overall costs at a wastewater treatment plant. Biosolids processing refers to the screening, grit removal, thickening, stabilization,

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_12 © Humana Press, New York, NY 2009

dewatering, drying, and disinfection of sludge, and also air emission control, if the target waste contains toxic volatile organic compounds (VOCs), and odorous substances.

This particular chapter introduces three new biosolids treatment processes, vertical shaft digestion (VSD), vertical shaft flotation (VSF) thickening, and gas-phase biofiltration. The combination of these three processes, and a few supplemental ones (such as grit removal, dewatering, and drying) provide complete biosolids treatment.

Specifically, this chapter discusses biosolids treatment objectives, theory and principles, description of processes (vertical shaft bioreactor [VSB], vertical shaft digestion, aerobic digestion, autothermal thermophilic aerobic digestion [ATAD], vertical shaft flotation thickening, optional anaerobic digestion, biosolids dewatering, and air emission control by biofiltration), engineering design, and case studies.

1.2. Vertical Shaft Bioreactor and Vertical Shaft Digestion

Vertical shaft bioreactor (VSB) is one of the advanced activated sludge processes for wastewater treatment (1, 2), while vertical shaft digestion (VSD) is one of advanced aerobic digestion processes for biosolids treatment (3, 4). Both VSB and VSD are alike from structural view points. Similarly, activated sludge process and aerobic digestion are similar to each other in terms of physical structure. Both VSB and VSD involve the use of vertical shaft reactors, which are typically 350 to 500 ft in depth, and 2.5 to 10 ft in diameter. Although vertical shaft reactors are usually constructed in ground, they can be constructed above ground when

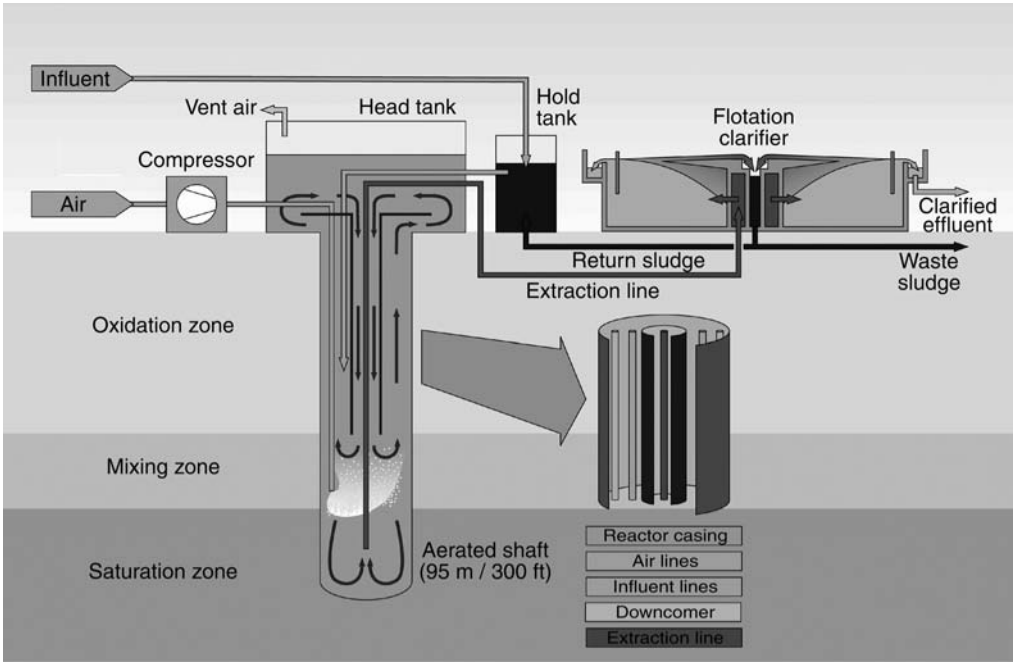


Fig. 12.1. Vertical shaft bioreactor (VERTREAT™) process flow diagram.

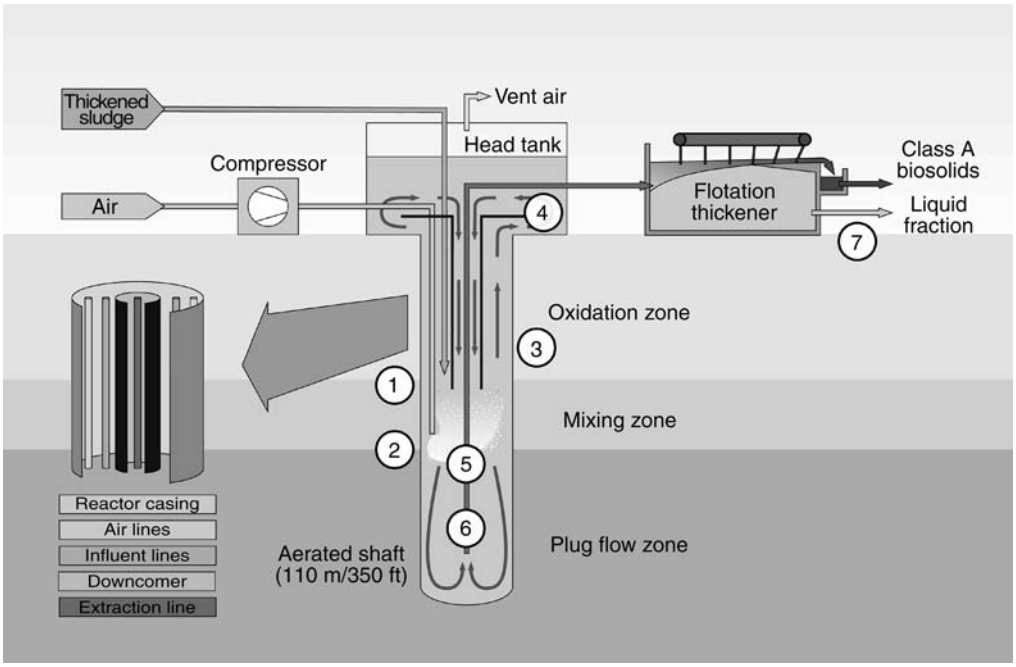


Fig. 12.2. Vertical shaft digestion (VERTAD™) process flow diagram.

necessary. Figures 12.1 and 12.2 show a VSB system for wastewater treatment and a VSD system for biosolids treatment, respectively.

When the vertical shaft reactor is used in a VSB system for wastewater treatment (Fig. 12.1), the reactor is mainly an aeration unit, and its influent is usually the primary wastewater effluent. The VSB effluent is treated by either a sedimentation or flotation clarifier, although the later is preferred. The clarifier effluent is then subjected to additional disinfection treatment, or advanced treatment (such as tertiary granular activated carbon adsorption, tertiary filtration, ion exchange, UV oxidation/disinfection, etc.) before its discharge to receiving waters. Typical commercial VSB processes include VERTRET and DEEP SHAFT, both of which have been extensively used in the UK, USA, and Japan. The readers are referred to other sources for detailed information on VSB processes (1, 2, 5–9).

When the vertical shaft reactor is used in a VSD system for biosolids treatment (Fig. 12.2), it becomes a vertical shaft autothermal thermophilic aerobic digester (VSD-ATAD) and its influent is usually a thickened sludge stream. A typical commercial VSD process is the VERTAD system manufactured by NORAM Engineering and Constructors, Ltd, Vancouver, Canada. A VSD system for biosolids treatment can be operated as a sequencing batch reactor (10, 11), or as a continuous biological digestion process. Either air or pure oxygen can be used for biosolids digestion/oxidation; therefore, VSD can be either a VSD-ATAD-Air, or a VSD-ATAD-Oxygen process. The VSD effluent is usually discharged to a flotation thickening unit and a dewatering unit for further biosolids–water separation. The dewatered biosolids are

either reused by spreading on agricultural land or sent to final land disposal (sanitary landfill) or for incineration.

It has been proven that the combined application of aerobic and anaerobic digestion will have a significant improvement upon the biosolids treatment efficiency. Accordingly, VSD has been frequently used for retrofitting the existing mesophilic anaerobic digesters. For this particular reason, the anaerobic digestion process and the results of the combined VSD-anaerobic digestion of biosolids are both briefly introduced in this chapter.

VSD is an advanced aerobic digestion process especially feasible for biosolids treatment in cold climate, where conventional aerobic digestion, or innovative cryophilic aerobic digestion (8, 9) are not cost-effective due to their big foot-print and above ground environment. VSD has extremely small foot-print, thus has small heat-loss. In addition, its vertical deep shaft reactor will allow the use of ground heat, and will exhibit a high oxygen transfer efficiency (OTE) due to the high hydraulic pressure (350 to 500 ft water column). VSD is also an attractive choice for biosolids digestion when the plant runs out of space for future expansion.

1.3. Vertical Shaft Flotation Thickening Process

Dissolved air flotation (DAF) is one of the best biosolids thickening processes (12). It is a well established technology and has widespread applications in wastewater treatment plants. Its only drawback is the high power cost for waste stream pressurization, gas injection and dissolution, gas release and micro-gas bubble generation.

Vertical shaft flotation (VSF) thickening process is always used in conjunction with the vertical shaft digester (VSD). The deep bioreactor serves as a pressurization tank for both gas injection and gas dissolution under high hydraulic pressure. The liquid biosolids stream in the bioreactor is supersaturated with dissolved gas. After the VSD effluent is discharged to a flotation tank under normal 1 atm pressure, the extremely fine gas bubbles will be generated due to sudden pressure release. The fine gas bubbles float the biosolids to the water surface forming a scum layer which is then skimmed off for further treatment (such as dewatering). The subnatant is recycled to the plant influent for liquid treatment.

The adoption of vertical shaft flotation saves:

1. Capital costs for pressure tanks and gas injection
2. Operation and maintenance costs for gas bubble generation
3. Treatment costs caused by sludge bulking problems

1.4. Gas-Phase Biofiltration

A liquid-phase biofiltration also known as trickling filter treatment has been used extensively for decades for wastewater treatment (13). Only recently, biofiltration was modified for treatment of air emission streams aiming at the reduction of biodegradable VOCs and odor-causing substances (14, 15).

Gas-phase biofiltration process equipment is commercially available (15). While the capital and O&M costs of gas-phase biofiltration are affordable, the gas stream collection cost is usually high. For this reason, conventional wastewater and biosolids treatment facilities usually do not have air emission collection and treatment systems installed.

A vertical shaft digestion (VSD) unit has a very small foot print (2.5 to 10 ft in diameter) making it feasible to install a complete air emission collection and biofiltration system for total environmental control. When wastewaters or biosolids are known to contain toxic biodegradable VOCs, one will seriously consider the inclusion of such equipment, a VSB for wastewater treatment, and/or a VSD for biosolids stabilization. Both systems can be totally covered for cost-effective air emission control using biofiltration.

1.5. Biosolids Digestion and Stabilization

Several techniques can be used for biosolids stabilization such as: anaerobic digestion, aerobic digestion, alkaline treatment, and composting. The primary purpose of stabilization is to reduce the biological activity of organic matter in the raw biosolids. Active organic matter can attract disease-carrying vectors such as flies. The secondary goals of stabilization are to reduce the mass of organic solids and the concentration of pathogenic bacteria. Given the multifaceted challenges that managers face, many are looking to advanced digestion to achieve their objectives related to:

1. Biosolids quality: reduce the pathogenic organisms and fecal coliform density
2. Solids reduction: reduce the amount of residual biosolids requiring hauling or tipping fees
3. Digester capacity: reduce the volume required for biosolids stabilization
4. Life cycle cost: reduce the life cycle cost of constructing, operating, and maintaining the digestion facility
5. Energy management: reduce the plant energy requirements
6. Operating characteristics: reduce odors, foaming, cleaning frequency, and impacts of side streams on wastewater treatment; improve mixing, heating, gas production for anaerobic digestion, and dewaterability

To achieve some of these objectives, we can look to high performance or advanced digestion processes. This chapter concentrates on vertical shaft digestion and related flotation thickening, dewatering, and optional anaerobic digestion processes. The key objectives relate to biosolids quality, solids reduction, digester capacity, life cycle cost, energy management, and various operating characteristics. The following subsections are based on the report of the Advanced Digestion Technology Team, Bioenergy Subcommittee, and Residuals and Biosolids Committee of Water Environment Federation (WEF) in 2002 (9).

1.5.1. Biosolids Quality

One of the driving forces for the development of advanced digestion technologies is the increasingly stringent regulatory requirements for biosolids handling. With increasing environmental regulation, rising disposal costs and a greater emphasis on public perception, the production of high quality biosolids is becoming more important. High quality biosolids have an improved public perception and a tremendous opportunity for beneficial use through land application (9).

High quality biosolids, as defined by 40CFR503, are divided into two classifications, Class B and Class A, based on the level of pathogen reduction achieved by the treatment process. All biosolids that are to be land applied for beneficial use must meet the requirements of one of these classifications.

Class B biosolids are usually achieved through a process to significantly reduce pathogens (PSRP), as defined by 40CFR503 (Appendix B). These processes include aerobic and anaerobic digestion, composting, lime stabilization, composting, and air drying. There are a number of restrictions on the harvesting of food crops, grazing of animals, and public access to land where Class B biosolids have been applied (9).

The production of higher quality, Class A biosolids offers the advantage of increased flexibility since there are few restrictions on the beneficial use or sale of Class A biosolids. In order to produce Class A biosolids, one of the seven alternatives listed in 40CFR503.32 must be met. The production of Class A biosolids through advanced digestion usually falls under Alternative 1, Alternative 5, and Alternative 6.

To meet the requirements of Alternative 1, sewage sludge is held at elevated temperatures for a specified amount of time, as determined by equations given in 40CFR503.32. To meet the requirement of Alternative 5, sewage sludge is treated in a process to further reduce pathogens (PFRP) as defined by 40CFR503 (Appendix B). Digestion coupled with pasteurization and thermophilic aerobic digestion are defined as PFRPs. Alternative 6 allows the regulating community to determine a given process to be equivalent to a PFRP (9).

1.5.2. Solids Reduction

Biosolids' solids reduction is one of the main objectives for sludge stabilization.

1.5.3. Digester Capacity

Some wastewater treatment plants have limited space available, but need greater digester capacity. Some advanced digestion technologies increase the capacity of existing tankage due to their shorter retention time and small foot print.

1.5.4. Life Cycle Cost

Whether an existing digestion facility is being upgraded or a new facility is being designed, minimizing life cycle cost is an objective.

1.5.5. Energy Management

Energy management involves demand-side management and resource management. Anaerobic digesters are being perceived more as an energy source than just a sludge stabilization technology.

1.5.6. Operating Characteristics

Operating characteristics include odor control, foam control, frequency of cleaning (surface and sediment). The complexity of a digester technology is an issue (9).

2. PRINCIPLES OF VSD AND OPTIONAL ANAEROBIC DIGESTION

2.1. Theory and Principles of Aerobic Digestion

Biological biosolids digestion is a method of sludge stabilization that uses bacteria to degrade organic matter. The principal purposes of stabilization are to make the treated biosolids less odorous, and to reduce the pathogenic organism content. Digestion also results in a substantial decrease in the mass of suspended biosolids.

There are seven different kinds of biological aerobic digestion processes (3, 4, 9, 16–29):

1. Conventional aerobic digestion using air (AD-Air)
2. Conventional aerobic digestion using oxygen (AD-Oxygen)
3. Conventional autothermal thermophilic aerobic digestion using air (ATAD-Air)
4. Conventional autothermal thermophilic aerobic digestion using oxygen (ATAD-Oxygen)
5. Vertical shaft digestion using air (VSD-ATAD-Air)
6. Vertical shaft digestion using oxygen (VSD-ATAD-Oxygen)
7. Cryophilic aerobic digestion

Vertical shaft digestion (VSD) is an autothermal thermophilic aerobic digestion (ATAD) process using either air or oxygen, therefore, VSD can be either VSD-ATAD-Air or VSD-ATAD-Oxygen process. Both VSD-ATAD-Air and VSD-ATAD-Oxygen use vertical shaft reactors (350 to 500 ft in depth), while both ATAD-Air and ATAD-Oxygen use conventional shallow bioreactors. The theory and principles of all aerobic digestion processes are alike and will be briefly covered in this section. For further details the readers are referred to another chapter (4).

The biological aerobic digestion process involves the direct oxidation of biodegradable matter and microbial cellular material by a biologically active mass of organisms. This is illustrated by the following reactions in the presence of microorganisms (13, 30–37):

$$\text{Organic matter} + O_2 \rightarrow \text{cellular matter} + CO_2 + H_2O \quad (1)$$

$$\text{Cellular matter} + O_2 \rightarrow \text{digested sludge} + CO_2 + H_2O \quad (2)$$

The second reaction (called endogenous respiration) is normally the predominant reaction in aerobic digestion (30–37). Endogenous respiration is the process whereby microorganisms metabolize their own protoplasm without replacement. Stabilization is not complete until there has been an extended period of primarily endogenous respiration (15 to 20 days for conventional aerobic digestion, and 4 to 6 days for vertical shaft digestion) if conventional aerobic digestion is applied. Vertical shaft digestion is operated under high pressure enhancing oxygen transfer and bio-oxidation., thus significantly reducing the required retention time for biosolids stabilization. Although the biochemical reactions shown in Eqs. (1) and (2) hold true for all aerobic digestion processes, the true mechanisms and kinetics of pressurized biochemical reactors, such as vertical shaft reactors are still unknown.

It is important to note that the nitrification (36) occurs in the mesophilic aerobic digestion processes (such as AD-Air and AD-Oxygen), but does not occur in the autothermal thermophilic aerobic digestion processes (such as, ATAD-Air, ATAD-Oxygen, VSD-ATAD-Air, and VSD-ATAD-Oxygen).

Small-scale aerobic digestion systems often use a one-tank sequencing batch system (11) with a complete mix cycle followed by settling and decanting (to help thicken the sludge). Larger operations may employ a separate sedimentation tank to allow continuous flow and facilitate decanting and thickening. Either air or pure oxygen can be used in these systems.

The aerobic digestion process is less sensitive to environmental factors than anaerobic digestion. However, one of its limitations is that it has less well established design parameters.

2.2. Theory and Principles of Optional Anaerobic Digestion

Vertical shaft digestion (VSD) occasionally is used to retrofit an existing failed mesophilic anaerobic digestion to form a combined system; therefore, the theory and principles of anaerobic digestion are also briefly covered in here (2, 4, 38–40).

Anaerobic digestion can be a single stage or a two stage digestion system. It can be operated as either a mesophilic anaerobic digestion or a thermophilic anaerobic digestion. The basic theory and principles of all anaerobic digestion processes are alike. Briefly speaking, anaerobic digestion is performed by several groups of anaerobic and facultative organisms that simultaneously assimilate and break down organic matter. It is a two-phase process. First, acid-forming organisms convert the organic substrate to volatile organic acids. Little change occurs in the total amount of organic material in the system, although some lowering of pH results. Alkaline buffering materials are also produced. Next, the volatile organic acids are converted primarily to methane and carbon dioxide.

This anaerobic process is essentially controlled by the methane-producing bacteria. These bacteria grow at a relatively slow rate and have generation times which range from slightly less than 2 days to about 22 days. Methane formers are very sensitive to pH, substrate composition, and temperature. If the pH drops below 6.0, methane formation ceases, and there is no decrease in organic content of the sludge. The methane bacteria are highly active in the mesophilic and thermophilic ranges. The mesophilic range is between 10°C and 47°C (50°F and 110°F) while the thermophilic range is between 45°C and 65°C (113°F and 149°F). Essentially, almost all digesters in the United States operate within the mesophilic temperature range.

Although very widely used, anaerobic digesters are sensitive to a variety of physical, chemical, and biological phenomena (e.g., pH, alkalinity, temperature, and concentrations of toxic substances). Anaerobic sludge digester biomass is relatively intolerant to changing environmental conditions. The process requires careful monitoring of pH, gas production, and volatile acids.

Anaerobic digestion can be performed in one or two stages. In single stage systems one tank is used for digestion and thickening. As decomposition proceeds, three distinct zones develop: the scum layer at the top of the digester, the supernatant zone in the middle and the sludge zone at the bottom. The sludge zones include an actively decomposing upper layer and a relatively stabilized bottom layer where the stabilized sludge accumulates. Two-stage anaerobic digestion evolved as an attempt to provide additional gas production as well as a separate settling and thickening process in the secondary digester. The readers are referred to another book of this handbook series for more detailed technical information on anaerobic digestion (41).

2.3. Combined Vertical Shaft Digestion and Anaerobic Digestion

Digestion reduces sludge volumes and produces less odorous biosolids that are often easier to dewater. Vertical shaft digestion (VSD) is an aerobic digestion process which has some advantages over anaerobic digestion including simplicity of operation, lower capital cost, fewer effects from interfering substances (such as heavy metals), and no danger of methane explosions. Since anaerobic digestion has a higher ability of reducing volatile solids content

than aerobic processes and since it has the advantage of producing methane as an energy source, VSD is occasionally applied in conjunction with anaerobic digestion (9).

Pilot testing is recommended before employing a combined aerobic–anaerobic biosolids digestion system to confirm and/or select the design and operating parameters.

The primary result of combined aerobic and anaerobic digestion system is the efficient reduction of volatile solids. The performance of aerobic digestion (such as VSD) depends on detention time, temperature, and character of solids. The performance of anaerobic digestion depends on proper seeding, pH, character of solids, temperature, and degree of mixing of raw solids with actively digesting seed material.

3. DESCRIPTION, OPERATION, AND APPLICATIONS OF VSD SYSTEM

3.1. Process Description

The principal difference between VSD and conventional ATAD systems is in its employing an in-ground hyperbaric reactor. Installed by conventional drilling techniques, the VSD reactor is typically 110 m (350 ft) deep, occupying only a fraction of the area used by conventional surface digestion systems. The diameter of the reactor, which can range from 0.75 to 3 m (2.5 to 10 ft) is determined by the quantity of biosolids requiring treatment. While traditional above-ground ATAD processes employ two or three tanks in series to achieve sufficient temperatures and prevent short-circuiting, VSD combines the stages within a single reactor.

As shown in Fig. 12.2, the VSD VERTAD™ reactor has three separate treatment zones: the oxidation zone, the mixing zone, and the lower plug-flow or soak zone. The oxidation zone is the upper portion of the reactor, and includes a central concentric draft tube for circulation. The mixing zone is immediately below the oxidation zone. Air required for bio-oxidation within the upper zone is injected into the mixing zone. The injected air also provides airlift circulation. The lower plug-flow zone is designed to prevent short circuiting and provides the high-temperature residence time required to kill pathogens such as salmonella and fecal coliform, ensuring that the product meets Class A biosolids requirements set forth by the US Environmental Protection Agency (US EPA) in CFR-503.

3.2. Process Operation

Figure 12.2 shows the flow diagram of a complete vertical shaft digestion system including supplemental processes for pretreatment, thickening, dewatering, and air emission control. The following shows how a vertical shaft digestion process is operated.

1. Screened sludge feed is delivered into the mixing zone where it is mixed with partially digested recirculating sludge.
2. Compressed air is continuously added below the mixing zone to provide the oxygen required by the microorganisms to digest the sludge. The high hydrostatic pressure ensures a high oxygen transfer rate (OTR).
3. Air bubbles rising up the outer annulus create circulation up the annulus, into the head tank, and down a central draft tube.
4. Off-gas containing excess air and carbon dioxide formed by microbial respiration disengages in the head tank and vents to an off-gas biofilter that effectively breaks any foam and removes odors.

5. A small fraction of the recirculating sludge moves from the mixing zone into the lower plug flow zone, which is designed to prevent short-circuiting. In this zone, residual organic materials are digested and the high temperature ensures that pathogens are destroyed.
6. Class A biosolids are withdrawn from the bottom of the reactor through a central discharge pipe and transferred rapidly to a product tank at the surface.
7. The rapid depressurization of the digested Class A biosolids causes the solids to separate in the product tank by flotation, and yields Class A biosolids pre-thickened to around 10% solids. The subnatant liquid is recycled back to the sewage treatment plant for processing prior to discharge.

3.3. Process Applications

As stated previously, vertical shaft digestion (VSD) is an advanced autothermal thermophilic aerobic digestion process. This technology employs a subsurface vertical reactor to aerobically digest mixed primary and secondary biosolids. Enhanced oxygen transfer in the process facilitates high metabolic activity resulting in heat generation. This enables the production of Class A biosolids at short solids retention times (SRT).

The VSD digestion system is commercially known as VERTAD, manufactured by NORAM Engineering and Constructors Ltd. (Vancouver, BC). Unlike conventional ATAD processes, the state-of-the-art VSD aerobic thermophilic process converts municipal primary and secondary sludges to Class A biosolids. It uses an in-ground hyperbaric aeration reactor—a device that has been proven effective through more than 20 years of commercial operation in biological treatment processes. The VSD reactor's patented design, according to its manufacturer, has the following advantages over conventional ATAD:

1. Excellent Volatile Solids destruction (>40% in a 4 day HRT).
2. Produces Class "A" biosolids product (40 CFR 503.32, Alternative 1).
3. Flotation thickening using dissolved gases in the product.
4. Thickened product dewaters to high solids content with low polymer demand.
5. Efficient space utilization due to its minimal plant footprint.
6. Highly efficient oxygen transfer.
7. Low volumes of process air to treat in subsequent off-gas biofilters.
8. Power costs are substantially lower than conventional aeration processes.
9. Enhanced microbial degradation due to efficient, high energy mixing.
10. Autothermal operation produces heat that is available for recovery.
11. Constructed using conventional well drilling or mining techniques.
12. Simple open-pipe aeration device requires very little maintenance.
13. Odor, VOC, and ammonia emissions are minimal compared to conventional processes.
14. Off-gas from head tank is contained and easily routed for biofilter treatment.
15. Lower capital cost than conventional Class A technologies.
16. The system can be economically enclosed in a building in locations where climatic conditions are unfavorable or if it is desirable for the plant to architecturally blend in with the surrounding environment.
17. The system is uncomplicated, easy to operate and maintain, and well suited to fully-automated unattended operation.
18. The in-ground reactor is much less likely to sustain damage in an earthquake than above-ground reactors.

The VSD system is ideal for treating biosolids streams from a VSB system, or from a conventional biological treatment plant treating municipal or industrial wastewaters. It has particular advantages in applications with the following conditions:

1. Sites with high biosolids disposal and/or trucking costs
2. Applications in which Class A biosolids are required
3. Sites with space constraints
4. Retrofits and plant expansions
5. Sites with high precipitation or extreme temperatures
6. Sites close to residential areas
7. Locations where large unsightly plants are undesirable (i.e., recreation areas)
8. Sites in areas with high seismic activity

4. DESIGN CONSIDERATIONS OF A COMPLETE VSD SYSTEM

A complete VSD system includes not only the main process digestion unit, but also the supplemental units, such as flotation thickener, supplemental anaerobic digester, biosolids dewatering unit, and air emission control system. The combination of the above units together will accomplish the objectives of biosolids stabilization, biosolids dewatering, supernatant recycling, and air emission control.

4.1. Autothermal Thermophilic Aerobic Digestion Using Air

Autothermal thermophilic aerobic digestion using air (ATAD-Air) is a form of advanced aerobic digestion that operates in the thermophilic temperature range (greater than 45°C) using air as the source of the required oxygen. The operation is autothermal that is; the heat required for the increase in temperature is supplied completely from the exothermic breakdown of organic and cellular material occurring during aerobic digestion. The increased temperature, in turn, reduces the required retention time for a given amount of solids reduction. The digesters are covered and insulated to minimize heat losses from the system. Use of oxygen in place of air (ATAD-Oxygen) is another similar advanced autothermal thermophilic aerobic digestion process, which is introduced in the next section. VSD can be either a VSD-ATAD-Air, or a VSD-ATAD-Oxygen, involving the use of vertical shaft reactor.

This section introduces only ATAD-Air and VSD-ATAD-Air. Both processes share the same theory and principles, except that VSD adopts a vertical shaft reactor instead of an above-ground tank. All design criteria developed for conventional ATAD-Air can be applied to VSD-Air as well. One full-scale ATAD-Air unit has been operated since 1977 at the Binghamton-Johnson City, New York wastewater treatment plant. Engineering results have fully demonstrated the feasibility of this process and have provided the technical know how presented below.

ATAD-Air and VSD-ATAD-Air can be applied to biosolids with solids concentrations of 1.5% or greater. More dilute biosolids will not reach thermophilic temperatures without supplemental heat. The high temperatures reached in the digester may result in virtually complete destruction of pathogens and eliminate the need for further disinfection. Thermophilic conditions can be reached in most climates and will require a much shorter retention time than unheated aerobic digestion or anaerobic digestion. At temperatures above 50°C, a high degree

of digestion and of solids reduction can be achieved with less than 8 days' retention. The high temperatures also decrease oxygen requirements because of the inhibition of nitrification. In general, aerobic digestion produces a supernatant with lower organic loadings than anaerobic digestion. The process may improve the settle-ability and dewatering characteristics of sludge. The simplicity of operation may be suitable for use in small treatment plants. It could also have application in cold climates where conventional aerobic digestion is ineffective or requires excessively long detention times.

The ATAD-Air process is not applicable to conventional waste activated sludges (WAS) because of the large amount of heat required to raise WAS (at 0.5% solids) to thermophilic temperatures. The process has high operating costs, primarily for air supply. The oxygen transfer efficiencies required to maintain thermophilic conditions with air may be as high as 15%. To achieve the high oxygen transfer efficiencies required, the system used was proprietary in nature; the "Liacom System" by DeLaval, Inc., which utilized a self-aspirating aerator. The VSD system (VARTRAD) marketed by NORAM, uses the deep shaft reactor as well as covers and jacketing to contain the heat.

Based on full scale ATAD-Air system studies, some selected parameters for a conventional (non-vertical shaft) 1000 ft^3 reactor are as follows (42):

1. Retention time, 5.4 to 7.7 day
2. TVS loading rate, 0.17 to 0.26 $lb/ft^3/day$
3. Treatment efficiency (% TVS removal), 22.1% to 37.2%
4. pH feed sludge, 5.4 to 6.1
5. pH reactor, 7.6 to 7.9
6. pH effluent, 7.6
7. Ambient temperature, 15°C to 25°C
8. Biosolids feed temperature, 20°C
9. Reactor temperature, 48°C to 52°C
10. Oxygen transfer efficiency, 8.7% to 15.1%
11. Air flow, 0.78 to 0.91 ft^3/s

Air adjustment, pH adjustment and mechanical foam cutting are generally required. Residuals generated include both the supernatant and the digested biosolids. General design criteria are: reactor temperature 45°C to 70°C and retention time 2 to 10 days.

The full-scale US EPA demonstration project indicated very few problems with the ATAD-Air process or equipment reliability. During winter conditions (ambient: −20°C) the digester remained in the thermophilic range. There were no operational problems with the self-aspirating aerator system. There are indications that the ATAD-Air process is generally more stable than anaerobic digestion and more easily able to recover from extreme conditions.

When vertical shaft reactors are used instead of conventional above-ground reactors, both oxygen transfer efficiency and treatment efficiency are higher and detention time is shorter. VSD-ATAD-Air requires less space than conventional digestion and, by stabilizing and disinfecting the biosolids, reduces the adverse impact of land application of biosolids.

4.2. Autothermal Thermophilic Digestion Using Pure Oxygen

Autothermal thermophilic aerobic digestion using pure oxygen (ATAD-Oxygen and VSD-ATAD-Oxygen) is a form of another advanced aerobic digestion that operates in the thermophilic (more than 45°C) temperature range and utilizes pure oxygen instead of air to aerate the sludge. The operation is autothermal, that is, the heat required for the increase in temperature is supplied completely from the exothermic breakdown of organic and cellular material occurring during aerobic digestion. The increased temperatures, in turn, reduce the required retention times in the digesters to achieve a given amount of solids reduction. The digesters are covered to minimize heat losses from the system. Heat loses are also reduced in pure oxygen systems because there is little exhaust gas to remove the heat generated by the process. The equipment for pure oxygen thermophilic aerobic digestion (ATAD-Oxygen and VSD-ATAD-Oxygen) is similar to that of the other advanced aerobic digestion (ATAD-Air and VSD-ATAD-Air) discussed previously with the addition of digester covers and an oxygen generator.

Two full scale studies (Denver, Colorado, and Speedway, Indiana) have been conducted using pure oxygen aerobic digestion since 1980. ATAD-Oxygen and VSD-ATAD-Oxygen systems may have greatest applications where pure oxygen activated sludge processes are used. The high temperatures used by the process may result in virtually complete destruction of pathogens, and eliminate the need for further disinfection. In colder climates the ATAD-Oxygen and VSD-ATAD-Oxygen processes will have much shorter retention times than other digestion processes. At temperatures above 45°C a high degree of digestion can be obtained with less than 5 days retention. The high temperatures decrease oxygen requirements because of the inhibition of nitrification. In general, all aerobic digestion processes produce supernatants with lower organic loadings than anaerobic digestion. The danger of methane explosions is also reduced.

ATAD-Oxygen process system may not be applicable to conventional un-thickened waste activated sludges (WAS) because of the large amount of heat required to raise WAS (at 0.5% solids) to thermophilic temperatures. The ATAD-Oxygen process has high operating costs (primarily to supply oxygen). No useful byproducts such as methane are produced. Oxygen aerobic digestion in the mesophilic temperature range does not appear to be cost effective, but in the thermophilic range the reduced requirements and smaller reactor volume may enable the process to be competitive with other forms of digestion, particularly when a pathogen-free sludge is desired.

Table 12.1 presents the US EPA's performance data for ATAD-Oxygen systems. The requirements of physical, chemical, and biological aids, and the generation of residuals of ATAD-Oxygen and VSD-ATAD-Oxygen systems are the same as those of ATAD-Air and VSD-ATAD-Air systems.

When vertical shaft reactors are used instead of conventional tank reactors, oxygen transfer efficiency and treatment efficiency will be higher, and detention time will be shorter. The design criteria for both single- and two-stage systems are similar: (a) retention time: 5 days or less, and (b) reactor temperature: 45°C to 60°C. The ATAD-Oxygen process, such as VSD, is stable and can more easily recover from extremes than anaerobic digestion.

Table 12.1
Performance of autothermal thermophilic aerobic digestion using oxygen (ATAD-Oxygen)

Single stage system	Phase I	Phase IA	Phase II	Phase III
Sludge description	O_2 step feed	O_2 step feed	O_2 activated sludge	primary +O_2 AS
Temperature (°C)	14–18	17–19	17.4–22	16–22
pH	6.0–6.3	5.9–6.4	5.9–6.4	5.5–6.1
TSS (mg/L)	25,000–33,000	30,000–34,000	25,000–40,000	–
VSS (mg/L)	21,000–27,000	22,000–27,000	20,000–30,000	–
TS (mg/L)	–	–	–	30,000–49,000
TVS (mg/L)	–	–	–	22,000–35,000
Retention time (days)	4.2	4.2	4.2	4.0
Digester temperature (°C)	47.3	46.4	50.4	50.2
VSS loading rate (lb/ft^3/day)	0.36	0.38	0.37	0.45
VSS reduction (%)	37	30	40	30

Two stage system – (multiple test runs combined)	O_2 waste activated sludge	Primary plus secondary sludge
Temperature (°C)	12–24	12–30
pH	5.9–6.9	6.0–6.6
TS (mg/L)	26,000–50,000	23,000–60,000
TVS (mg/L)	18,000–38,000	18,000–41,000
Retention time (days)	3.7–5.0	3–5
Digester temperature (°C)	48.7–57.8	45.3–52.0
VS loading rate (lb/ft^3/day)	0.32–0.46	0.38–0.53
Overall VSS reduction (%)	29–42	30–45

4.3. Flotation Thickening after Vertical Shaft Digestion

In conventional dissolved air flotation (DAF) systems, recycled subnatant flow is pressurized to 30 to 70 lb/in.^{2} gage and then saturated with air in a pressure tank. The pressurized effluent is then mixed with the influent sludge and subsequently released into the flotation tank (12).

The flotation thickener after vertical shaft digestion (VSD), however, is not a conventional DAF, because the VSD effluent has already been pressurized in the reactor to a high pressure (350 ft of hydraulic water head). The VSD effluent containing supersaturated gas can be directly released in a flotation tank for biosolids thickening. This new flotation process is called vertical shaft flotation (VSF), a process that does not need the conventional DAF pressure tank (gas dissolving tank, or gas dissolving tube). The excess dissolved gas in the VSD effluent separates from solution at the atmospheric pressure in the VSF thickener. The minute, 80 μm, rising gas bubbles attach themselves to biosolids particles which form the floating sludge blanket at the water surface. The floating thickened biosolids are skimmed off and pumped to the downstream biosolids handling facilities while the subnatant is returned to the plant's headworks. Polyelectrolytes are frequently used as flotation aids to enhance performance and create a thicker biosolids blanket. A flow diagram of the VSD-VSF process system is shown in Fig. 12.3.

Although VSF is a new technology, yet it is similar to DAF in efficiency, theory, and principles. DAF is the most common form in the United States for the thickening of waste activated sludges, and to a lesser degree combined sludges. DAF also has widespread indus-

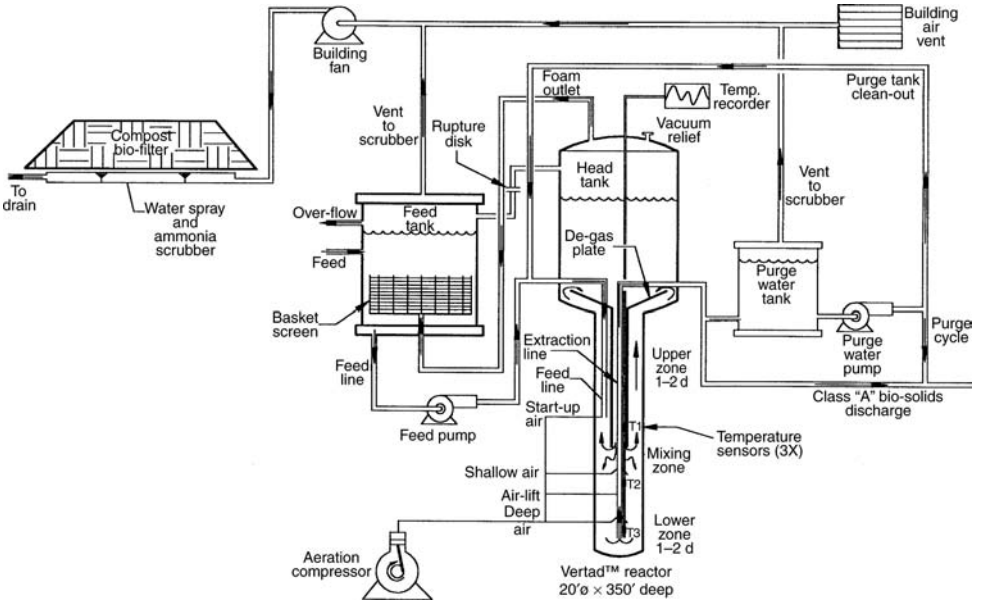


Fig. 12.3. Vertical shaft digestion (VERTAD™) demonstration facility process flow diagram.

trial wastewater applications. It is expected that VSF will compete with DAF favorably from both technical and economical view points.

The use of vertical shaft flotation is limited primarily to the thickening of biosolids following vertical shaft digestion and prior to dewatering or anaerobic digestion. Used in this way, the efficiency of the subsequent dewatering units can be increased and the volume of resulting supernatant is decreased. Existing vertical shaft flotation thickening units can be upgraded by the optimization of process variables and the utilization of polyelectrolytes.

With VSF thickening, it is possible to attain biosolids concentrations of up to 6% compared to a maximum of 2% to 3% that can be achieved for WAS in gravity thickening. Data from various flotation thickening units indicates that solids recovery ranges between 83% and 99% at solids loading rates of 7 to 48 lb/ft^2/day. Flotation aids, mostly polyelectrolytes, are commonly used to enhance performance.

VSF thickening requires less land area than gravity thickeners. The subnatant stream is returned to the head of the treatment plant. The gas released to the atmosphere may strip volatile organic material from the biosolids. The volume of sludge requiring ultimate disposal may be reduced, although its composition will be altered if chemical flotation aids are used.

4.4. Optional Dual Digestion System

The process of dual digestion involves the use of an aerobic digestion process as a pretreatment step before mesophilic anaerobic digestion. Dual digestion is a well established Class A biosolids process. The majority of current systems utilize an aerobic digestion step with a short contact time (around 24 h) and pure oxygen to support the biological process (28). There are two possible approaches to dual digestion that involve operating the aerobic first

stage at different retention times. In one scheme the aerobic stage is operated at an HRT of 1 to 2 days to achieve Class A pathogen removal and some level of stabilization (10% to 20% VS removal). A second approach involves a longer HRT in aerobic digestion (between 4 and 6 days) which would achieve Class A pathogen removal and a more significant level of stabilization (35% to 50% VS removal).

In VSD systems (either VSD-ATAD-Air or VSD-ATAD-Oxygen), organic nitrogen and FOG are preferentially degraded over organic solids comprised primarily of cellulose. This is significant when considering a dual digestion flow sheet with VSD pretreatment ahead of anaerobic digestion. The technologies are complementary in that the VSD systems can readily degrades fats and proteins, compounds known to cause scum build-up and mixing problems in mesophilic anaerobic digesters, and the anaerobic digestion process is capable of destroying the cellulose material typically present in VSD product (3).

Using a dual digestion system will result in increased overall volatile solids destruction compared to that of either VSD or anaerobic digestion alone. Systems employing dual digestion can achieve as high as 70% VS destruction in a 15 day SRT (3, 27). This compares favorably to VSD systems which achieve a VS destruction of 40% in a 4 day SRT, and mesophilic anaerobic digestion which can achieve 50% VS destruction in a 20 day SRT.

Although the VSD process does not generate methane gas, it does produce recoverable heat in the form of hot water, which can be used to heat the mesophilic digesters in a dual digestion system. The methane produced in anaerobic digestion can then be used for other purposes.

The majority of POTWs in North America still utilize mesophilic digestion as the sole process for stabilizing sludge. This process successfully produces methane gas as well as a stabilized Class B biosolids product. The impetus for facilities to explore combined digestion is the fact that many of these municipal treatment facilities generate significant quantities of biological solids, and need to maximize solids destruction in order to minimize solids handling and disposal costs (3). Retrofitting an existing mesophilic anaerobic digestion system to a dual digestion system offers the following benefits:

1. Production of Class A Biosolids
2. Increased overall volatile solids destruction (as high as 70% VS destruction in a 15 day SRT compared to 50% VS destruction in a 20 day SRT in mesophilic digestion alone)
3. Heat recovered from the ATAD process can be utilized for sludge conditioning, as well as building and anaerobic digester heating
4. Methane is produced in anaerobic digestion, and the overall bioenergy recovery from the dual digestion system is higher than that from aerobic or anaerobic alone
5. Improved operation in the mesophilic anaerobic digestion stage (improved mixing, less scum, operation at higher solids concentrations)
6. Significantly reduces the size (or increases the capacity) of the dewatering system, improved dewaterability
7. Reduces recycle nutrient loading to the treatment facility

4.5. Biosolids Dewatering Processes

Dewatering is the removal of water from biosolids to achieve a volume reduction greater than that achieved by thickening. Dewatering of biosolids is desirable for one or more of the following reasons:

1. To prepare sludge for landfilling
2. To reduce sludge volume and mass for lower transportation costs
3. To reduce the moisture content and thereby increase the net heating value to make incineration more economical

Some dewatering processes use natural means, such as evaporation, percolation, etc. (43) for moisture removal; others use mechanical devices to speed the process. The method chosen for dewatering is determined mainly by the type of biosolids, space available, subsequent processes, and economics.

The most common biosolids dewatering methods are:

1. Vacuum filtration
2. Filter press
3. Belt filter
4. Centrifugation
5. Thermal drying
6. Drying beds
7. Lagoons

All of these technologies are well established. The most commonly used methods for dewatering industrial biosolids are lagoons and drying beds. The mechanical methods for dewatering biosolids using vacuum filtration, centrifuges, and filter presses are also in widespread use.

All biosolids dewatering processes except drying beds and lagoons are complex mechanical systems. Their reliability is thus dependent on operator skill and proper maintenance. Vacuum filtration requires considerable operating attention and proper chemical conditioning to prevent filter blinding. Filter and belt filter presses have several moving parts and require maintenance to obtain a high level of reliability. Centrifuges are high speed mechanical devices subject to maintenance problems.

The performance of dewatering devices is measured by biosolids concentration or cake moisture, and solids recovery. Dewatered concentrations of 10% to 30% solids are common with biosolids and values of 60% solids or more may be attained with some inorganic residues. The performance of any one specific dewatering method depends on biosolids type, characteristics, conditioning, and operating conditions.

4.6. Gas-Phase Biofiltration for Air Emission Control

A vertical shaft digestion system is similar to other aerobic digestion processes; therefore, an air emission control unit for removal of odor and volatile organic compounds (VOCs) is not absolutely required. Occasionally the digester influent does contain toxic VOCs and odorous substances, and the aerobic digester has to be enclosed for air emission control. Under this adverse condition where enclosure is required, conventional aerobic digesters and

conventional autothermal thermophilic aerobic digesters will not be economically feasible due to their big foot prints. The construction costs of an enclosure for collection and subsequent treatment of polluted air emission streams will be extremely high. Vertical shaft digesters with typical diameters in the range of 2.5 to 10 ft, would have small foot prints that will allow VSDs to be economically enclosed for air emission. This section introduces one of many air emission control technologies which can be applied in conjunction with vertical shaft digestion units. Biofiltration frequently teams up with vertical shaft digestion due to its low cost. Air emission control technologies and their related costs can be found in "*Air Pollution Control Engineering*" and "*Advanced Air and Noise Pollution Control*" Refs. (15, 44).

4.6.1. Biofiltration Process Description

Biofiltration is an emerging technology for controlling volatile organic compounds (VOCs) emission in waste gas streams. Biofiltration has been extensively used in Europe, especially for odor control and it has been demonstrated at full-scale in the United States (14). In the biofiltration process, the waste gas is vented through a biologically active material where the biodegradable VOCs are oxidized into carbon dioxide and water. Physical sorption and chemical degradation may also occur and contribute to the overall removal efficiency. Figure 12.4 is a schematic diagram of a typical single-bed biofilter system. Since biofilters are biologically sensitive, the temperature and moisture of the gas and filter bed are extremely important in design considerations. Radial blowers are used to transport the waste gas to a humidifier. The humidifier saturates the gas stream to 95% relative humidity, which prevents drying out of the filter material. The effect of the filter drying out is death of the microorganisms and a resultant loss of control efficiency.

The gas stream enters the gas distribution system below the filter. As the gas diffuses through the filter, air contaminants will diffuse into the wet, biologically active layer (biofilm) where degradation occurs. Clean gas diffuses out the top of the filter. Excess drainage from the filter bed is the only potential source of wastewater discharge. In particular, where drainage contains regulated organic contaminants, the drainage is recycled to the humidifier to minimize wastewater discharge. Since particulates in the waste stream may clog the humidifier and

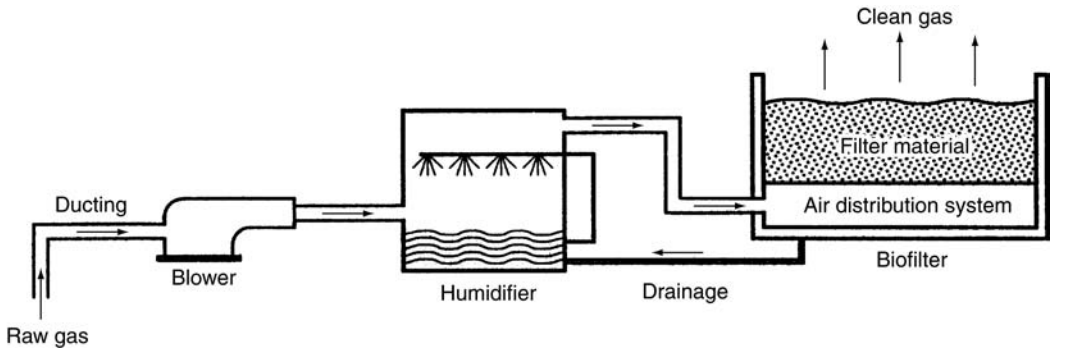


Fig. 12.4. Biofiltration flow diagram.

the biofilter, a pie-filter may be required. A heat exchanger may also be required to heat or cool the waste gas stream if temperatures are not within the optimum range (20°C to 40°C).

Typically, the filter material is compost, peat, wood chips, or soil with an inert material such as polystyrene particles. As the VOCs are degraded, water, carbon dioxide, mineral salts and biomass are generated. Mineralization leads to compaction of the filter material, which causes an increase in backpressure. Typically the filter material is turned over after 2 years of operation and usually replaced 1 to 2 years after turning over the filter to prevent backpressure problems (45).

The most common biofilter system is an open, single-bed system. The clean gas is vented directly to the atmosphere in an open biofilter. Enclosed, multiple-bed systems can be stacked and have been employed for low maintenance and space constraint situations.

4.6.2. Applicability to Air Emission Control

The applicability of biofiltration is dependent on the characteristics of the waste gas emitted from a (VSD) or other treatment process. Typical biodegradable contaminants include: alcohols, ethers, aldehydes, ketones, amines, sulfides, and certain monocyclic aromatics (xylene, benzene, toluene, and phenol). Waste streams containing chlorinated solvents are not readily biodegradable and are not appropriate for emissions control by biofiltration.

Biofiltration, as a VOC control technology, results in the complete degradation of the biodegradable contaminants and avoids the cross media transfer of pollutants. A major requirement, and thus limitation, of biofiltration is the absence of biologically toxic substances in the waste gas, such as heavy metals. The technology is limited to biodegradable components.

Since biofiltration is biologically sensitive, the potential system failures represent areas that should be considered when evaluating this technology. An undersized filter can result in VOC air emissions due to insufficient treatment. Since the filter is sized by off-gas flow rate and concentration, the off-gas should remain within these design parameters during operation to prevent the loss of control efficiency. Inadequate preconditioning of the off-gas for temperature, moisture, particulates, or toxic constituents can also result in the complete loss of control efficiency.

Intermittent off-gas streams can be treated with a biofilter assuming the flow rate and concentration of the gas stream are within the design values. Filter beds can survive shut down periods of at least 2 weeks without any significant reduction in biological activity. Shut down periods up to 2 months are feasible with nutrient addition and aeration of the filter (14, 45).

Biofiltration is technically and economically feasible for controlling VOCs in large volume gas streams with low concentrations. One potential use of biofiltration is odor control at POTW sites assuming the odor constituents are biodegradable. Since odor problems usually are caused by compounds with low odor thresholds, off-gas concentrations often will be relatively low.

4.6.3. Range of Effectiveness

Biofiltration usually is cost effective for large volume gas streams with relatively low concentrations (<1000 ppm as methane) of easily biodegradable contaminants (14). Maximum

influent VOC concentrations have been found to be 3000 to 5000 mg/m^3 (45). For optimum efficiency, the waste gas should be 20°C to 40°C and 95% relative humidity. The filter material should remain at 40% to 60% moisture by weight and have a pH between 7 and 8. For most easily biodegradable constituents, control efficiencies greater than 90% are achievable (14). Degradation rates for common air pollutants are typically from 10 to 100 g/m^3-h (45). The key parameters affecting the control efficiency of a biofiltration system include the environmental conditions in the filter material, biofilter design, filter size, and waste gas composition. The filter must also have a large reactive area and low pressure drops; therefore, compaction must be kept to a minimum.

4.6.4. Sizing Criteria of Biofiltration

Typical biofilter systems have been designed to treat 1000 to 150,000 m^3/h waste gas with the systems having 10 to 2000 m^2 of filter area (14). The depth of biofilter material is typically 3 to 4 ft. The size of a biofilter system is dependent on the following parameters:

(a) The loading rate of waste gas
(b) The concentration of compounds in the waste gas
(c) The rate of degradation of the compounds per unit volume

Surface loads up to 300 m^3/h of waste gas/m^2 filter area are feasible without excessively high back pressures (45). The type of filter material affects the pressure drop across the filter. The effect of filter material on pressure drop is shown in Fig. 12.5 as a function of the surface loading rate.

The filter's large mass often provides sufficient buffer capacity to prevent breakthroughs during peak loadings, which allows sizing based on average hourly peak loads (14).

The removal process in biofilters has been postulated to be controlled initially by a first-order-type biodegradation rate, but to be limited by transport properties at low inlet air flow rates (14). Pilot testing of industrial waste gas streams with multiple contaminants is usually required, rather than modeling, to accurately size the full scale system.

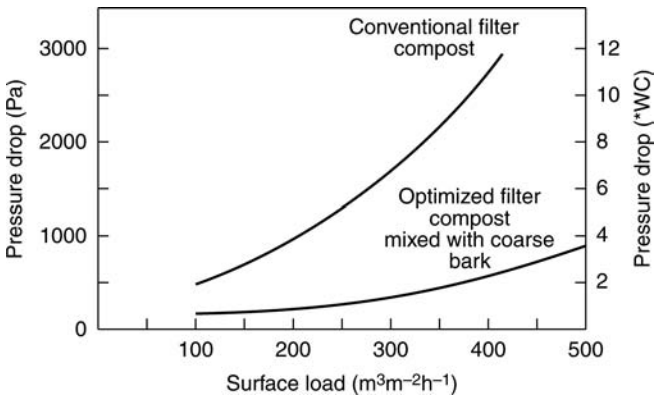


Fig. 12.5. Pressure drop of two biofiltration systems as a function of surface loading rate.

4.6.5. Cost Estimating Procedure

Capital costs have been estimated at 77 to 123 USD/ft^2 filter area for installed open, single-bed biofilter systems. Costs of open, multiple-bed systems are approximately two times these costs. Enclosed systems have been estimated to cost between 123 and 677 USD/ft^2 filter area, depending on the size of the biofilter and the degree of process control (45). Operating costs are 0.45 to 2.06 USD/100, 000 standard ft^3, not including filter replacement costs (14). Maintenance costs are about 1 USD/m^2 of filter/year.

Cost estimates were updated from 1991 to reflect the 2007 costs using the Cost Index for Utilities (Appendix A); all costs were multiplied by a factor of 539.74/392.35=1.38 (46).

4.7. Operational Controls of Biofiltration

Operational controls are those procedures or practices inherent to the operation (and design) of control systems that can be followed to minimize the overall long-term emissions. Among these are:

1. Adequate system design and installation
2. Startup testing
3. Preparation of standard operating procedures (SOPs) for operators
4. Control of operating variables to minimize emissions
5. Monitoring of system performance
6. Minimization of process upsets and startups, and
7. Preventative and routine maintenance

Obviously, a properly designed and operated control system is necessary to achieve the required air emission control efficiency or air emission limits. The use of experienced contractors and vendors will help ensure that the system design and installation are done correctly. Startup testing is advisable, with as many test conditions examined, as possible and all meaningful data should be recorded and evaluated. Systematic checks of wiring, direction of fan and pump rotation, integrity (leak tightness), etc. should be made. The startup testing results should be incorporated into the formal standard operating procedures (SOPs) prepared for and followed by the operators of the biofilter.

Operating variables can be controlled to minimize air emissions. The most obvious variable to control is the treatment rate; e.g., the lower the feed rate to additional air emission equipment, the lower the mass of potential emissions. Other variables such as the aeration rate for biodegradation systems, also directly influence emissions. Controlling operating variables to minimize air emissions is not always straightforward. There may be a number of competing variables that must be balanced for optimal control system performance.

To properly operate control devices, the biofiltration system design and performance must be understood. Performance data can be generated by routine monitoring of influent and effluent emission levels, pressure drops, operating temperatures, and so on. Operators should maintain the monitoring system so that plugged lines, water in the lines, etc. don't result in misleading readings. Proper maintenance is another obvious requirement for successful control system operation, including routine inspection of the equipment and implementation of corrective action when needed.

5. CASE STUDY

A demonstration project was supported by the Technology Assessment Program of the King County (WA) Wastewater Treatment Division (47). This program was developed in 1991 to evaluate and test technologies to reduce the environmental impacts of treatment plant operations including the space required for solids handling, biosolids truck traffic and odor emissions. The vertical shaft digestion technology was selected for evaluation based on the potential for a very small footprint, low odor emissions, and production of Class A biosolids. A demonstration facility was constructed in 1998 and operated through 1999 at the South Treatment Plant (STP) in Renton, WA. Successful results in these tests have prompted the County to consider VSD as a retrofit for existing facilities, and for future projects (3).

The VSD demonstration project consisted of design, construction, and operation of a demonstration-scale, deep vertical reactor for thermophilic aerobic digestion located at the South Treatment Plant (Renton, WA). The project team led by E&A Environmental Consultants, Inc. (E&A) was responsible for the planning, design, construction, and testing of the facility. The technology owner, NORAM Engineering and Constructors Ltd., was actively involved in all aspects of the test program. King County provided engineering, operations, and maintenance support throughout the project. The facility was completed in January, 1998 and testing was completed in December, 1999.

The test program was based on the following objectives:

1. To evaluate the SRT and temperature requirement for compliance with the Vector Attraction Reduction and Class A pathogen requirements of the US EPA 40 CFR 503.32 Alternative 1
2. Evaluate reactor hydraulics, oxygen transfer efficiency (OTE), and energy balance
3. Determine the dewaterability of the VSD effluent (cake solids, polymer demand)
4. Evaluate "dual-digestion"—VSD as pretreatment for mesophilic anaerobic digestion
5. Perform an economic analysis of the technology

5.1. Facility Design and Construction

The demonstration facility is located at the South Treatment Plant (STP) operated by King County in Renton, WA. The wastewater treatment plant is a 115 MGD facility with primary clarification, activated sludge secondary treatment, co-thickening of primary and secondary solids by dissolved air flotation (DAF), anaerobic digestion and belt press dewatering (3, 47). A summary of the design parameters for the demonstration facility is provided in Table 12.2.

The main component of the VSD facility is a 50 cm diameter, 107 m deep (20 in. × 350 ft) subsurface, vertical reactor. The reactor tube was placed by conventional drilling technology using the dual air rotary drilling method. Subsurface geology consisted of 50 m of coarse sand and gravel alluvium above bedrock of siltstone and shale. There were indications of flowing groundwater above the bedrock. Prior to project initiation, the County conducted an assessment of the potential for earthquake damage to a deep reactor. The study concluded that damage to the reactor likely would be less than that to surface tankage (48). This finding is consistent with similar studies that have been carried out for Pacific Rim installations including Japan, Alaska, British Columbia, and California.

Table 12.2
Design parameters for vertical shaft digestion demonstration facility

Influent characteristics	Value
Influent	Thickened municipal biosolids (THS)
Loading	500–1500 lb solids/day (2500–7500 pop. equivalent)
Solids concentration	6.5%
VSS	78% to 80%
Primary sludge	60%
WAS	40%
Temperature	20°C to 21°C
Biofilter treatment building exhaust	800 cfm
Biofilter loading rate	5 cfm/ft^2
Feed rate	
@ 3 day HRT	1770 gpd
@ 6 day HRT	889 gpd
Equipment Inventory	
VERTAD™ BIOREACTOR	
Casing	1 @ 20 in. diameter by 350 ft deep
Draft tube	1 @ 10 in. diameter by 143 ft deep
Extraction	1 @ 3 in. diameter by 347.5 ft deep
Reactor volume	
Total	740 ft^3
Liquid	710 ft^3
VESSELS	
Feed tank	1 @ 60 in. diameter by 72 in. high
Digester head tank	1 @ 60 in. diameter by 72 in. high
Purge water tank	1 @ 38 in. diameter by 48 in. high
MECHANICAL	
Aeration compressor	87 scfm, 150 psi, 25 HP
Feed pump	1–10 gpm, 50 psi, 3 HP
Purge water pump	20 gpm, 26 ft TDH, 5 HP

The vertical reactor has three separate treatment zones. A diagram of the process illustrating these three treatment zones is shown in Fig. 12.3. The upper zone of the shaft (surface to 44 m depth) contains a central concentric draft tube for circulation. The shallow aeration header introduces compressed air below the draft tube to induce flow up the annular space and down the draft tube. Thickened solids are introduced into this completely mixed zone. The lower zone extends below the draft tube down to the deep aeration header (44 to 96 m depth). High oxygen transfer rates are attained in this zone under pressures of 5 to 10 atm. Mixing between the upper and lower zones occurs gradually over several hours. An unaerated plug-flow zone extends below the deep aeration header to the bottom of the reactor (96 to 107 m depth). This zone is hydraulically separate from the aerated upper zones (as confirmed by tracer tests).

Stabilized product is withdrawn using airlift through a 7.6 cm pipe that extends to within 0.5 m of the bottom of the reactor. The effluent is batch discharged at intervals sufficient to ensure strict adherence to the time/temperature requirements for pathogen destruction of Class A biosolids.

The support equipment for the reactor includes a thickened solids (THS) supply loop, a feed storage tank, a feed pump with variable frequency control, a purge water system, a 25 hp air compressor, a heat exchange system, a programmable logic controller (PLC), and a biofilter for off-gas treatment. The batch effluent withdrawal and feeding cycles (continuous or batch) are fully controlled by via PLC. Levels are continuously monitored by differential pressure sensors in the feed tank and reactor head tank. Temperatures are continuously monitored by sensors hanging at five elevations in a wet well in the center of the reactor. The THS supply loop provides a continuous supply of fresh undigested solids from the STP solids system storage tank.

The feed tank provides 2.2 m^3 of feed storage. Process air to the reactor is provided by a continuous duty, rotary screw compressor that requires 18.6 kW to produce 2.5 m^3/min at 1035 kPa (87 scfm at 150 psi). Compressed air is injected at 48 and 96 m depths. The process air supplies oxygen for biological metabolism and induces mixing in the reactor. Air that is not dissolved produces voidage (volume of bubbles per unit volume of liquid) and is released from the reactor liquor in the head tank. A weighted check valve on the off-gas pipe provides up to 35 kPa (5 psi) back pressure to the head tank which reduces voidage. The off-gas is directed to the bottom of the feed tank to provide additional back pressure and capture foam and latent heat in the influent solids. The test facility is housed in a temporary building that is provided with utilities and air collection. Building exhaust and reactor process off-gas passes through a water scrubber for ammonia removal and is then processed through a biofilter.

A system to add supplemental heat to the reactor was installed after it became evident that heat loss from the pilot reactor exceeded the heat generated biologically and thermophilic temperatures could not be maintained. The reactor was not insulated and has a high surface area to volume ratio, which facilitates heat transfer to the environment. Also, flowing water was identified in three zones during drilling. Water moving past the reactor can remove substantial heat. To compensate for heat loss to the environment, reactor feed was initially preheated via steam injection using an 80,000 Btu/h propane-fired steam boiler. This was replaced later in the test period with a boiler (500,000 Btu/h) that supplied hot water to heat exchanger loops hanging in the reactor. This system provided direct control of the temperature in the reactor. The supplemental heating system was added rather than using a sludge–sludge heat exchanger to capture heat from the effluent. While the VSD plant referenced is often cited as a pilot or demonstration scale facility, it should be noted that this plant can process the solids from a 7500 population equivalent. A 7500 population equivalent would be serviced by a 0.75 MGD VSD facility that would feed solids to a VSD plant of roughly this size. So while this plant is considered small by King County standards, it would be a full-scale facility for smaller municipalities.

Table 12.3
Summary of operating ranges

Operating variable	Operating range
Hydraulic residence time (day)	2–6
Temperature (°C)	55–70
Aeration (scfm)	20–80
Feed solids content (%)	3.5–7

5.2. Vertical Shaft Digestion Demonstration Plan

The vertical shaft digestion demonstration program was designed to meet the goals of the King County research program to evaluate the viability of the technology with respect to reactor hydraulics, energy requirements, product quality and the ability to meet the vector attraction reduction and pathogen destruction requirements of Class A biosolids. An additional goal was to develop the design criteria necessary for full-scale design and cost evaluation.

A range of operating conditions was tested in the facility. Prior to biological startup, cold water testing was conducted to evaluate reactor hydraulics and to test equipment. Next, pre-heating of the reactor using a hot water boiler provided data on heat loss to the environment in the absence of biological heat generation. Biological testing with varied HRT, temperature, aeration rates, and feed solids spanned the periods of January 15 to May 7, 1998, November 10 to December 17, 1998, and August 4 to December 23, 1999. Suspensions in operation between the various testing periods allowed ongoing modifications of the facility for improved data acquisition and control. During the third testing period, stable operation was achieved over a range of detention times and temperatures. In order to test a full range of conditions and determine the capabilities of the VSD process, some operating conditions were applied that did not provide a Class A biosolids product. However, these imperative tests provided insights into the critical effects of such variables as sludge viscosity, oxygen transfer efficiency, and heat loss. The range of operating conditions that were tested for the process is summarized in Table 12.3.

The controlled parameters of the test program were the solids retention time (SRT), temperature, and aeration rate. The approach was to establish stable operations at specified operating conditions for a minimum of three detention times. During the fourth detention time, data was averaged to yield the reported values for the reactor performance under those stable conditions. Samples were collected for laboratory analysis of the thickened solids feed (THS), feed and head tank (upper zone) solids, and final effluent solids (from the deep extraction line). These samples were tested for total solids (TS), volatile solids (VS), pH, total Kjeldahl nitrogen (TKN), ammonia, chemical oxygen demand (COD), and alkalinity (ALK) by the STP laboratory according to Standard Methods. Fecal coliform and salmonella analyses were conducted by the King County Environmental Laboratory. Additional laboratory and field testing included measurement of fat, oil and grease (FOG), total carbon (TC), total organic carbon (TOC), off-gas analysis, density testing, oxidation reduction potential (ORP), dewaterability, and dissolved oxygen (DO). Daily grab sample analyses of TS and VS were

conducted while the remaining parameters were measured weekly. More frequent sampling and composite analyses were conducted during the fourth detention time. Temperatures, levels, and flows were logged and trended continuously via PLC using a Siemens WinCC trending program. Oxygen concentration in the off-gas was measured using a portable oxygen analyzer (first with a Quintox gas analyzer, and later a Teledyne Portable Flue Gas Oxygen Analyzer). The dewatering characteristics (polymer demand, cake solids content, and filtrate quality) of the digested product were tested by several dewatering equipment vendors (CIBA, US Filter, Andritz). Five gallon samples were delivered to vendor laboratories where testing was performed on bench scale centrifuges, belt presses, and capillary suction time (CST) test equipment. Onsite testing was conducted to compare VSD product to the mesophilically digested STP biosolids using jar testing to determine polymer demand, and press tests to assess the maximum achievable dry cake solids content.

Tracer tests using both lithium chloride (LiCl) and table salt (NaCl) were conducted to assess the reactor hydraulic characteristics and to confirm that no short-circuiting was occurring in the reactor. LiCl or NaCl was batch loaded into the reactor and samples were collected from the reactor head tank and the product during batch product withdrawals. In the case of tracer tests involving LiCl, the samples collected were analyzed for Lithium content, and profiles were developed. For the salt traces (only performed in water), conductivity changes were measured in the samples taken from several depths in the reactor. Samples from lines at 213 and 268 ft depths were drawn continuously by a peristaltic pump at a rate of 1.3 L/min through 3/8″ ID tubing weighted to keep it in place. Enough salt was added to increase the conductivity to approximately ten times the background concentration, ensuring good resolution. Conductivity was measured using the STP conductivity analyzer after proper temperature equilibration. This has automatic temperature compensation so readings need no further correction. Bench-scale testing was conducted at the University of Washington to assess the effect of VSD pretreatment on subsequent mesophilic anaerobic digestion (dual digestion). VSD product (4 day SRT) was fed to 3L anaerobic digesters maintained at 11 and 15 day detention times. A control digester was fed STP thickened solids at an 11 day SRT. The digesters were maintained at 35°C. The main parameters used to evaluate digester performance included volatile solids destruction efficiency, gas production and percent methane, and product dewaterability using CST testing.

Odor panel testing was performed on samples collected from the VSD process. The odor panel analyses were conducted by Odor Science & Engineering, Inc (OS&E). These tests were aimed at measuring the odors generated by the VSD process and the effectiveness of the biofilter for odor treatment. Odor was quantified by dilution-to-threshold (D/T) ratio and panelists described the odor character.

5.3. Design Criteria Development for Vertical Shaft Digestion

5.3.1. Volatile Solids Destruction

A summary of the digestion performance results is presented in Table 12.4. The values reported are averages over a detention time after the process was stable for three detention

Table 12.4
Volatile solids reduction at varied temperature and residence times

Test	HRT (day)	Temperature (°C)	Aeration rate (scfm)	VS reduction (%)
Dec'98	4	56	56	40.9
Sept'99	4	65	80	42.2
Nov'99	3.4	56	36	42.3
Dec'99	5.5	61	30	43.5

times. A complete mass balance was achieved for each of these tests from which the reported efficiency values were calculated (3).

The effect of solids residence time on VS reduction was demonstrated by the testing. Greater than 40% VS reduction was demonstrated at a 4 day SRT. This efficiency appears to decrease approximately linearly as the residence time is reduced. In testing at a 2 day SRT and 67°C, a 21% VS reduction was demonstrated. As shown in Table 12.4, a 5.5 day SRT resulted in a 43.5% VS reduction. This value is considered conservative because concurrent testing of reactor response to oil and sugar addition complicated results due to the additional load on the system. Results from the three detention time conditioning period for the 5.5 day SRT test were averaging at 50.7% VS reduction prior to the supplemental additions. From these results it is believed that VS reduction will approach 50% with a detention time of approximately 6 days at 60°C (360°C-day).

In general, it was found that an increase in temperature for a given solids retention time resulted in greater VS reduction. Testing indicated that although temperature certainly affects biological activity, it is believed that the effects on water loss and oxygen transfer efficiency on reactor performance are much more significant and important. Important findings about the effect of reactor sludge viscosity on oxygen transfer resulted in testing centered on controlling the reactor solids. With solids controlled at below 4.5% TS, oxygen transfer efficiency nearly doubled, allowing a subsequent decrease in aeration rates. Decreased aeration rates minimized the amount of water loss (as latent heat) from the reactor for a given temperature. The difference between the Dec'98 and Nov'99 results can be explained by this finding. The two test periods were both operated at a temperature of 56°C, however the Nov'99 trial was operated at a reduced SRT (3.4 days compared to 4 days), and at a reduced aeration rate (36 scfm compared to 56 scfm). The major difference between the two trials was that in the case of the Nov'99 trial, the reactor solids concentration was being controlled at 3.5% TS, and in the Dec'98 trial, the reactor solids concentration was 4.7%. Ultimately, the increased ability to transfer oxygen in to the mixture allowed a decreased SRT while simultaneously provided increased VS reduction.

The requirements for Class A biosolids were met at an average detention time of 4-days at 60°C. As shown in Table 12.3, the system readily achieved greater than 40% volatile solids destruction at varied detention times and temperatures. In order to satisfy the volatile solids destruction criteria of 38% (24) in conventional ATAD systems, Kelly et al. (29) suggested a

400°C-day product was necessary. The VSD results indicate that a 240°C-day product exceed the US EPA requirements, with greater than 40% volatile solids destruction.

5.3.2. Pathogen Destruction

Pathogen destruction was excellent with a 7 log reduction in fecal coliform and both fecal coliform and salmonella below detection limits in the Class A biosolids product (3). Fecal coliform and salmonella were measured in the feed solids and digested VSD effluent weekly during the first operating period and intermittently during the third operating period. Fecal coliform in the feed solids averaged 5.39E + 07 MPN/g dry solids and salmonella averaged 5.87 MPN/4 g dry solids. Densities in the VSD effluent were consistently below the detection limit (fecal coliform: 5 MPN/g, salmonella: 1.6 MPN/4 g).

5.3.3. Reactor Mixing

The selected alternative for attaining Class A pathogen control in the VSD process is by maintaining temperatures for the required contact time. Time and temperature requirements from the biosolids regulations (40 CFR 503) are shown in Fig. 12.6.

In order to test the reactor's compliance with time-temperature requirements, salt tracer studies were performed in the system. Samples were taken at regular intervals from four points in the system: the head tank (surface), 213 ft below grade surface (bgs), 268.5 ft bgs,

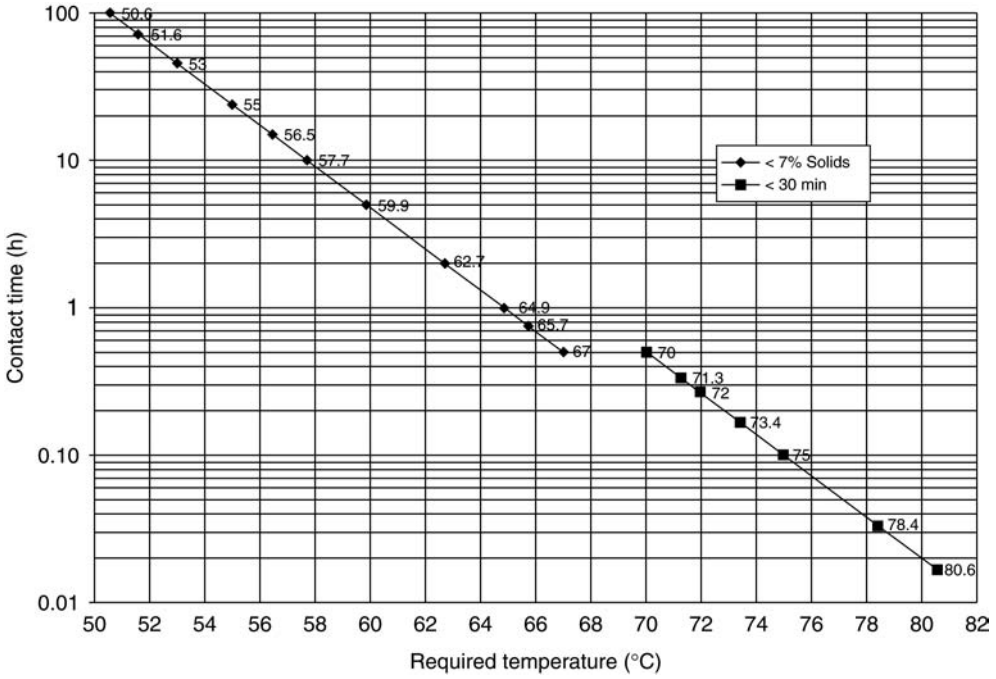


Fig. 12.6. US EPA 40 CFR 503 class a time and temperature requirements for solids less than 7%.

and the deep extraction line. Critical distances in the system are: Upper Aeration Head—158 ft bgs, Lower Aeration Head—315 ft bgs, and Deep Extraction Line—347 ft bgs. Head tank and intermediate sample points allowed observation of the saline dispersion as it moved through the reactor, providing an indication of the mixing time between the aeration headers. The deep extraction point allowed observation of the saline pulse, showing the time for a single particle to breakthrough the soak zone. A pulse of saline was pumped into the reactor quickly with enough salt for a tenfold increase in reactor conductivity. After the pulse of saline, the system was fed and discharged continuously at a rate of approximately 2 gpm (HRT of approximately 2 days). The conductivity profile versus time for the reactor tracer study is shown in Fig. 12.7 (3).

Tracer results are consistent with a model in which:

1. The upper zone (head tank to upper aeration head) is well mixed, with a time constant of the order of minutes.
2. The lower zone (upper aeration head to lower aeration head) is mixed gently by fluid rising in the wake of bubbles with a net turnover time which depends strongly upon air flow. In this study the lower aeration was 8 scfm, resulting in gentle mixing over approximately 90 min. Here, simple theory based on the assumption that a bubble draws up its own volume of fluid are in reasonable agreement.
3. The soak zone is effectively plug flow.

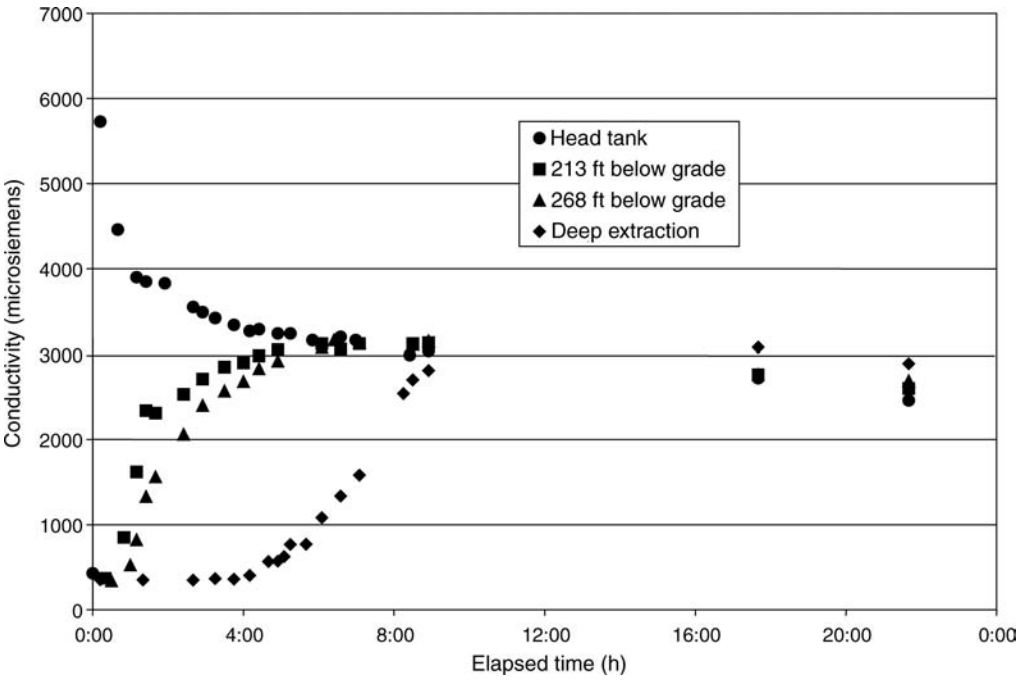


Fig. 12.7. Vertical shaft digestion salt tracer study confirming no short circuiting.

The mixing test clearly indicates that the salt tracer did not reach the deep extraction point until approximately 4 h had elapsed. This eliminates any concerns about short-circuiting in the reactor soak zone. The theoretical time for breakthrough (based on the 2 gpm extraction rate and the soak zone volume for plug flow) is 4 h 20 min. This is the first continuous feed, single reactor design that complies with the US EPA time-temperature regulations. Salt tracer studies confirmed that the VSD patented reactor design complies with time-temperature requirements (40 CFR 503 Class A Time and Temperature Requirements for Solids Less Than 7%). These studies verify the true plug flow nature of the soak zone, and eliminate any concerns about short-circuiting in the system. While it is believed that the demonstration facility's vertically stacked zone configuration complies with the time and temperature requirements, two variations are available to further assure compliance: (a) Installation of a flow restricting physical barrier between the slowly mixed and soak zones, and (b) Maintaining a surface batch contact tank in which the VSD product is held for the required time at the appropriate temperature (using heat generated from the VSD).

5.3.4. Vertical Shaft Flotation Thickening

During dewatering testing, it was indicated that the VSD effluent could be easily thickened after being discharged from the reactor. VSD effluent has the characteristic of high dissolved carbon dioxide concentrations due to the biological metabolism and the high pressure in the reactor. Acidifying the effluent (with sulfuric acid or alum) to approximately a pH 5, releases the CO_2 as small bubbles, which attach to biosolids particles and float them to the compact surface blanket. Further testing resulted in float thickening of the VSD biosolids from 3.5% TS to 8% to 12% total solids, with a capture efficiency of approximately 95%. Results were similar with both sulfuric acid and alum. Ferric chloride was not used, but it is expected to provide a similar result (3).

Analysis of the float thickened solids and the subnatant showed that nutrients partitioned into the digested biosolids. Thickened biosolids contained a phosphorus concentration 20 to 40 times the concentration in the subnatant. In the tests where sulfuric acid was used, the formation of ammonium sulfate caused the ammonia to get slightly partitioned into the biosolids. This result means that the phosphorus load that is typically recycled to the secondary treatment plant is being retained in the biosolids for beneficial reuse. The downstream implications of this flotation thickening step are as follows:

(a) Significantly reduction in the size of the dewatering system
(b) Charge neutralization aids in dewatering
(c) Reduced recycle nutrient loading on the treatment facility
(d) Increased nutrient value of the Biosolids

5.3.5. Biosolids Dewatering

Test methods for dewatering included onsite press tests as well as outside laboratory testing at Andritz, IBA, and other vendors using bench scale belt presses and centrifuges. Samples tested included mesophilic anaerobic sludge from the STP, biosolids directly from the VSD reactor, VSD float thickened biosolids, and product from the combined VSD to anaerobic bench-scale test work (3, 47).

Table 12.5
Onsite press testing of vertical shaft digestion product dewaterability

Characteristics	Anaerobic	VERTAD™	Acid float thickened VERTAD™
Cake solids (%)	20	32	31
Polymer dose (lb/t)	17	17	17
Filtrate quality	Clear	Very turbid	Very clear

Table 12.6
Andritz lab centrifuge testing of vertical shaft digestion product dewaterability

Characteristics	Anaerobic	VERTAD™	Acid float thickened VERTAD™
Cake solids (%)	12–14	31–34	31–34
Polymer dose (lb/t)	20.4	38	13.8
Capture efficiency (%)	95	96	99.5

Onsite press testing was performed using a set polymer dose of 17 lb/t for the mesophilic anaerobic sludge from the STP, biosolids directly from the VSD reactor, and the VSD float thickened biosolids. Cake solids were measured and the filtrate quality was reported qualitatively. The results are presented in Table 12.5.

Testing demonstrated that greater than 30% cake solids could be attained with both the biosolids directly from the VSD reactor and the VSD float thickened biosolids whereas the anaerobically digested solids dewatered to 20% cake solids. In the case of the VSD reactor biosolids, the filtrate quality was poor, with losses of solids making the filtrate look very turbid. This indicated that a higher polymer dose would be required with the straight VSD product to obtain an acceptable filtrate quality. The VSD float thickened product outperformed both the anaerobic and VSD products. Not only did the VSD float thickened product have a very clear filtrate (clearer than that from the anaerobic dewatering tests); it obtained the best result from a cake solids perspective. This testing illustrated that the float thickening process greatly enhances the dewaterability of VSD biosolids. Outside laboratory testing at Andritz, CIBA, and other vendors, was performed on samples of mesophilic anaerobic sludge from the STP, biosolids directly from the VSD reactor, and the VSD float thickened biosolids. Polymer dosing was optimized using 95% solids capture efficiency for the filtrate quality. Cake solids and solids capture efficiency were measured and reported. The results are presented in Table 12.6.

The Andritz test results showed that greater than 30% cake solids could be attained with both the VSD reactor product and the acid float thickened product. Similar to onsite press testing, higher polymer doses were required for the VSD product withdrawn directly from the reactor (approximately double the polymer required for the anaerobic sludge). The anaerobic

biosolids dewatered very poorly with the lab centrifuge, only attaining a maximum cake solids concentration of 14% (3). Like onsite press testing, the VSD acid float thickened product showed remarkable dewatering characteristics. It dewatered to high cake solids concentration (31% to 34%) with a lower polymer dose than that required for anaerobic sludge (13.8 and 20.4 lb/t, respectively). The conclusion is that the float thickening enhances the dewaterability of the VSD product. This is likely due to a charge neutralization that seems to act like a coagulant, aiding in dewatering. It is generally accepted that thermophilically digested aerobic biosolids can be dewatered to higher cake solids than anaerobically digested biosolids; however this has historically come with the expense of greater polymer demand. Murthy et al. (26) performed an examination of an autothermal process to isolate the cause of high polymer demand and high recycle chemical oxygen demand (COD). They found that the presence of monovalent ions in solution such as sodium, potassium, and ammonium ions can interfere with charge-bridging mechanisms occurring in the floc. This is a problem in conventional ATAD systems because the release of ammonium ions is the result of the absence of nitrification in the thermophilic process (25). This free ammonia release appears to be less pronounced in the VSD process, possibly due to the pressure in the reactor which results in the combination of free ammonia with dissolved CO_2, forming ammonium bicarbonate.

Murthy et al. (26) also found that the amount of biopolymer (proteins and polysaccharides) in solution was heavily correlated to increased polymer demand. They concluded that the concentration of biopolymers in solution was minimized by limiting the solids retention time (SRT) of thermophilic digestion, and by minimizing the concentration of monovalent ions (specifically ammonia) in solution. These factors favor the VSD process because a relatively short SRT of 240°C-day is enabled by the high oxygen transfer achieved in the system, and ammonium bicarbonate is formed in the reactor, minimizing free ammonia.

5.3.6. Organic Nitrogen and FOG Destruction

A summary of the digestion performance results for VS, FOG, and organic nitrogen is presented in Table 12.7. The values reported are averages over one detention time after the process was stable for three detention times. A complete mass balance was achieved for each of these tests from which the reported efficiency values were calculated.

The reduction of organic nitrogen and fats, oils, and grease were relatively high considering the short solids retention times that the VSD process was tested at. The results were similar to the reduction efficiencies attained in the STP anaerobic digesters at a 28 day SRT. The organic

Table 12.7
Volatile solids, FOG, and organic nitrogen reduction

Test	HRT (day)	Temperature (°C)	VS reduction (%)	Org-N reduction (%)	FOG reduction (%)
Dec'98	4	56	40.9	57.9	91.7
Sept'99	4	65	42.2	49.8	80.8
Nov'99	3.4	56	42.3	44.1	–
Dec'99	5.5	61	43.5*	49.9	80.4

nitrogen reduction was calculated based on the difference between TKN and ammonia in the feed and product. Organic nitrogen reduction generally exceeded the total VS reduction. Analysis of the Dec'98 samples showed that protein degradation (assuming 6.25 kg protein/kg org-N) and FOG reduction accounted for 64% and 9%, respectively, of the VS reduction. The remaining VS reduction was attributed to carbohydrate reduction which is primarily comprised of cellulose and lignin. The preferential degradation of Org-N and FOG was further confirmed by visual inspection of the product which is very fibrous. These results are significant since undigested Org-N and FOG are generally responsible for the objectionable character of biosolids. These results also have significance when considering a dual digestion flow sheet with VSD pretreatment ahead of anaerobic digestion. The technologies appear to be complementary in that the VSD technology readily degrades fats and proteins, compounds known to cause scum buildup and mixing problems in anaerobic digesters, and the anaerobic digestion process is capable of destroying the cellulose material still present in the VSD product.

5.3.7. Biofiltration for Odor and Off-Gas Control

In the VSD system, the self-contained nature of the head works allows easy control over off-gas emissions. Off-gas can be easily routed to biofilters to remove the trace ammonia and dimethyl sulfide (DMS) compounds common with aerobic digestion technologies. Because of the high oxygen transfer efficiency in the bioreactor, the VSD process needs only a fraction of the air volume used in a conventional ATAD. As a result, significantly less off-gas is produced in the VSD process, reducing the size of biofilter required for off-gas treatment. Gaseous emissions from the VSD system are considerably smaller than those produced in conventional aeration processes. As mentioned previously, ammonia is converted to ammonium bicarbonate in the reactor, helping to eliminate ammonia emissions. In order to minimize the ammonia release from the system, reactors are operated at a maximum temperature of 60°C, preventing the dissociation of the ammonium bicarbonate.

Odor panel testing was performed on samples collected from the VSD process. The odor panel analyses were conducted by Odor Science & Engineering, Inc (OS&E). These tests measured the odors generated by the VSD process and the effectiveness of the biofilter for odor treatment. Odor was quantified by dilution-to-threshold (D/T) ratio and panelists described the odor character. The results of the odor panel work are provided in Fig. 12.8.

These results show that most of the VSD demonstration facility derived odor comes from the feed tank (16,463 D/T in 675 scfm) rather than the VSD reactor (1468 D/T in 36 scfm). The biofilter removed 99.5% of the odor loading (16,463 D/T in, and 79 D/T out). Odor panel testing has indicated that the off-gas from the VSD process is generally odor-free. Character descriptors for the VSD off-gas prior to the feed tank included more terms such as compost, earthy, and vegetation. The off-gas from the untreated feed sludge tank changed the odor panel characterizations to focus on terms such as sludge and manure type odors. These results have highlighted the need to treat the off-gas directly from the reactor in a biofilter.

The reduced odor of the VSD off-gas is primarily attributed to the fact that the compounds primarily responsible for the objectionable character of unstabilized wastewater solids (FOG, Org-N) are the highest degraded fractions in VSD.

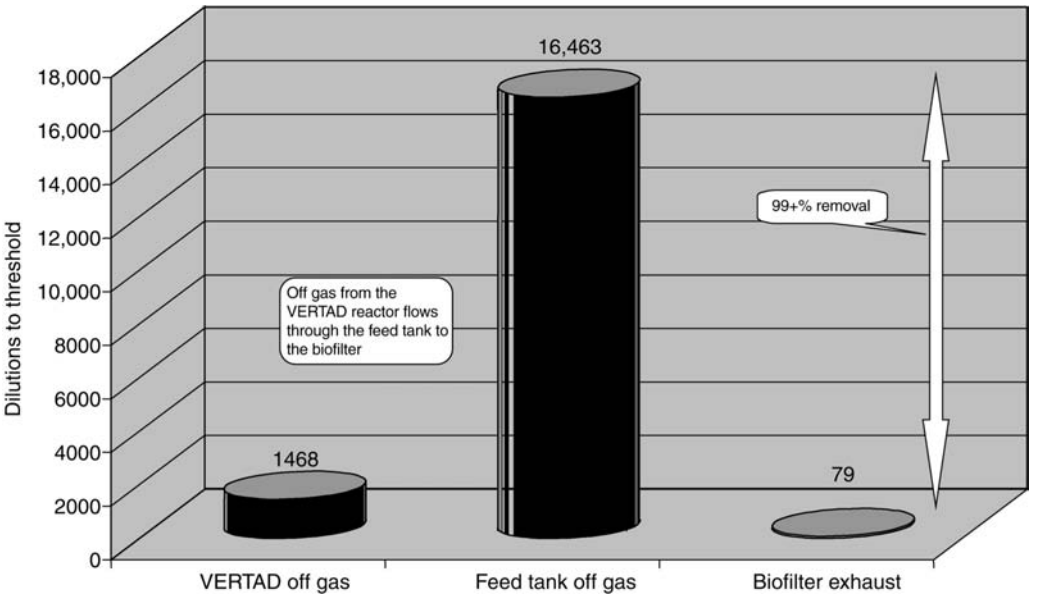


Fig. 12.8. Odor panel results of biofiltration process.

5.3.8. Oxygen Transfer Efficiency

Oxygen transfer studies were performed to test the oxygen transfer rate (OTR) first into water and determine the theoretical maximum efficiency of the system, and second into sludge to determine the oxygen transfer efficiency (OTE) attainable in the digestion process. The test method used to determine the OTR into water came from the ASCE (American Society of Civil Engineers). This test involved the initial scavenging of dissolved oxygen with sodium sulfite and a cobalt chloride catalyst (Na_2SO_3 and $CoCl_2$), followed by reoxygenation to near the saturation level for the operating temperature. Throughout these tests DO was measured at multiple points, allowing the development of a mass transfer model. The OTR was measured for water, allowing calculation of the OTE in the system. The OTE was approximately 66% into water at 54°C (129°F). This OTE represents a significant advancement in aeration technology over conventional aeration systems using air, which typically attain 10% to 20% OTE into water at 20°C, a lower temperature which facilitates oxygen transfer through increased solubility (3).

Sludge viscosity was found to have a pronounced effect on the OTE. As shown in Fig. 12.9, an OTE of 50% was attained easily at a reactor concentration less than 4.5% TS. At greater than 4.5% TS, the transfer efficiency was diminished, as low as 35%.

Although sludge is highly non-Newtonian, and the concept of a Newtonian viscosity which is independent of shear rate is not strictly valid, some valuable order of magnitude generalizations can be made about transfer performance at higher VS destruction. In general, the OTE is improved at higher VS destruction because the viscosity of the bulk liquid is decreased with increased destruction, and decreased viscosity facilitates increased oxygen

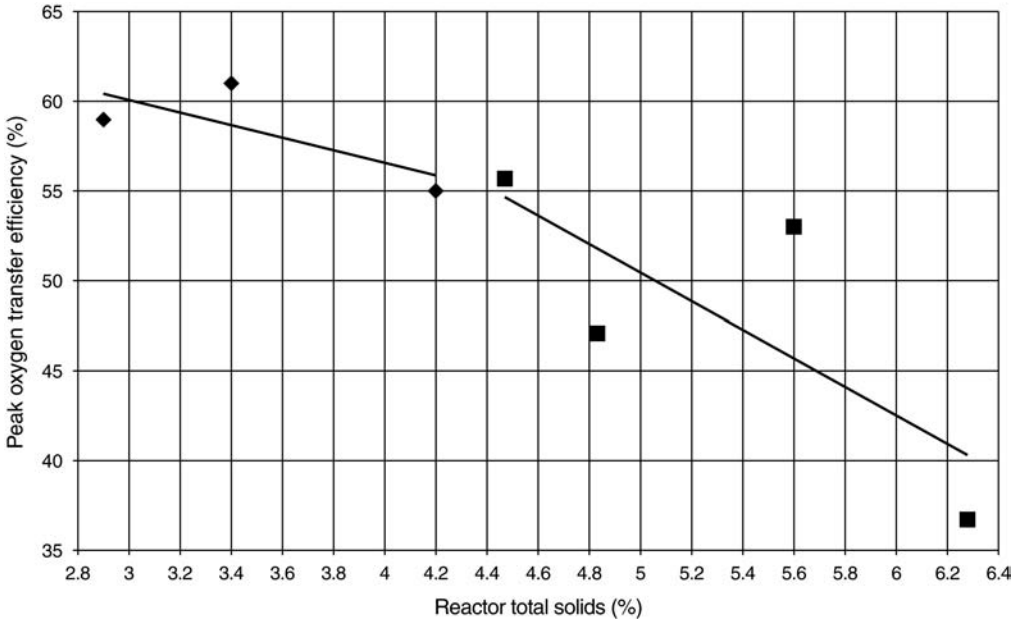


Fig. 12.9. Viscosity effects on peak oxygen transfer.

transfer. Transferring oxygen into thick sludge is not easy—even at high pressure, due to mass transfer limitations on the liquid side. Metcalf and Eddy (28) suggest that viscosity may decrease by a factor of 2 or more over the range of 3% to 6% for undigested sludge, with viscosity declining rapidly as sludge is digested. Doubling fluid viscosity changes oxygen diffusivity in the sludge in inverse proportion (i.e., it is halved); the mass transfer coefficient and mass transfer rate will likely change by a similar order of magnitude. This is supported by the VSD OTE data which suggests that the OTE is nearly halved with a doubling in reactor solids, and that 4.5% is the practical operating cutoff before the sludge viscosity seriously affects the OTE. The effect of oxygen transfer upon heat release was corroborated during the oxygen transfer testing. During testing each test involving a lower aeration rate saw a systematic decrease in the reactor temperature. Each time the aeration rate was reduced, biological heat generation was reduced and a step change in temperature occurred.

The VSD system achieves very high oxygen transfer efficiency, greater than 50% OTE can be expected when the viscosity of the reactor contents is controlled with a solids concentration less than 4.5% TS. These high oxygen transfer rates are associated with the pressure and depth at which compressed air is introduced to the bioreactor. The high OTE results in enhanced digestion of the sludge and a decreased detention time to meet the Class A biosolids requirements.

The OTE for other ATAD systems is generally not reported in literature presumably due to the proprietary nature of the systems, however, some independent data collected from an ATAD facility suggests that the VERTAD™ process compares favorably in terms of

oxygen transfer efficiency. While the VERTAD™ system achieves an average oxygen transfer efficiency of 50%, a conventional ATAD system that was tested only achieved an average of approximately 24% across a three stage system, presumably due to the high viscosity and low temperatures in early stages.

The increased oxygen transfer in the VSD system is thought to be the primary factor in decreasing the solids retention time to meet US EPA vector attraction requirements. As mentioned previously, Kelly et al. (29) have suggested that a 400°C-day product is necessary in ATADs to attain a volatile solids destruction of 38%. The VSD results indicate that a 240°C-day product exceed the US EPA requirements, with greater than 40% volatile solids destruction. The difference in oxygen transfer and subsequent heat release in the two systems could explain this superiority of the VSD system.

5.3.9. Heat Balance

The small diameter of the demonstration reactor results in a large surface area to volume ratio, necessitating supplemental heat addition at the facility to maintain the required elevated temperature. A heat balance was performed using measured reactor heat loss data, influent and effluent temperatures, estimated biological heat production, aeration energy and the measured supplemental heat necessary to maintain a set temperature. The heat balance showed that autothermophilic conditions would be maintained if the reactor diameter were increased to 0.8 m. (2.6 ft), thus decreasing the relative surface area. Reactors of larger diameter will require a heat removal system to prevent overheating, and recovered hot water will be available to the treatment plant for space heating and for digester heating in linked anaerobic systems.

5.3.10. Vertical Shaft Digestion Process Simplicity and Stability

The biological process was found throughout the testing program to be relatively simple to operate, resistant to upset, and to rapidly recover from disruptions caused by electrical and mechanical system failures. The straightforward process controls consist of providing a supply of food on a relatively uniform basis and providing air. In a full-scale system the operational controls are expected to require less operator attention than in an anaerobic digestion process. The VSD process operates well over a range of pH conditions and temperatures. Although the process does not generate gas, it does produce hot water and does not require the extensive gas handling, cleaning, and safety equipment.

The ability of the process to recover quickly from upset conditions was demonstrated on numerous occasions as the result of power outages and failure at the feed system, boiler, or control system. During these occasions, the process was stressed by lack of food, cooling, and aeration. In all situations the process recovered rapidly.

5.3.11. Vertical Shaft Digestion Followed by Anaerobic Digestion (Dual Digestion)

The process of dual digestion involves the use of an autothermal aerobic digestion process as a pretreatment step before mesophilic anaerobic digestion. In conventional dual digestion systems the aerobic step usually has a contact time of about 24 h and pure oxygen is typically used to support biological metabolism. Dual digestion is a well established Class A biosolids

Table 12.8
Dual digestion using vertical shaft digestion and mesophilic anaerobic digestion

Comparison of Anaerobic Control with Combined System Performance			
	11 day SRT Anaerobic control	15 day Anaerobic with VERTAD™	11 day Anaerobic with VERTAD™
Solids retention time (day)			
VERTAD™	0	4	4
Anaerobic	11	15	11
Total	11	19	15
Volatile solids reduction (%)			
VERTAD™	0	40	40
Anaerobic	52	49	45
Total	52	70	67
Anaerobic gas production			
Methane (L/day)	2.8	2.0	2.5
Methane (L/g COD removed)	0.51	0.39	0.36

process. The VSD process was evaluated as a pretreatment step to mesophilic anaerobic digestion. The impetus to test the combined digestion is the fact that King County treatment facilities, and many biosolids generators, need to maximize solids destruction in order to minimize solids handling costs. The effect of VSD pretreatment on subsequent mesophilic anaerobic digestion was tested using bench scale reactors in studies performed at the University of Washington by Jenny Yoo (3). The results of the dual digestion study are presented in Table 12.8.

The results indicate that following VSD with mesophilic anaerobic digestion provides additional reduction of volatile solids with the production of significant gas volume. Anaerobic digestion of the VSD product resulted in 67% total volatile destruction with a 4-day SRT in VSD and an 11-day mesophilic anaerobic SRT, and 70% total volatile destruction with a 4-day SRT in VSD and a 15-day mesophilic anaerobic SRT. Comparatively, a control anaerobic digester obtained 52% VS destruction with an 11-day SRT. While the control digester showed greater VS reduction in the anaerobic stage than the VSD fed anaerobic digesters (presumably due to the lower VS content in the feed from VSD), the total reduction efficiencies of the dual digestion systems were much higher than that of the anaerobic control.

The technologies appear to have a synergy from a performance and operability perspective. For example, the VSD technology readily degrades fats and proteins, whereas anaerobic digestion is capable of cellulose destruction. Observations during the bench scale testing were that the control digester experienced considerable foaming and had mixing problems. The dual digestion systems had no foaming problems and were readily mixed, indicating lower viscosity. This may be attributed to the efficient Org-N and FOG destruction in the VSD process. The ability to float thicken the VSD effluent presents itself as another benefit for the

combined system. Thickened product could be fed to anaerobic digestion, allowing operation at higher solids concentrations. The lower volumetric flow associated with the thicker feed would allow for either reduced digester volume requirement or increased solids retention time.

Biosolids with higher total solids concentration would decrease the volumetric flow to dewatering equipment and would likely improve dewatering performance. Several high solids concentration processes are currently being advocated including the anoxic gas flotation process (49). Qualitative indications from the limited dewatering testing of the combined product were that it dewatered to high cake solids (estimated at 24% cake solids) at very low polymer doses (5 to 6 lb/t).

Incorporation of a post-VSD mesophilic anaerobic digestion step shows considerable promise. The technologies appear to be complementary in many respects from a solids destruction and operability standpoint. This synergy of technologies results in enhanced VS destruction (up to 70%) making VSD an attractive retrofit option for existing anaerobic systems. The minimal footprint requirement for the VSD process make it an ideal retrofit for facilities that require additional capacity, or current Class B biosolids generators that wish to produce Class A biosolids. Figure 12.10 is a schematic showing the South Treatment Plant with a VSD retrofit that could either pre-treat the entire sludge stream to Class A time-temperature criteria (similar to Concept 2 from Fig. 12.11) or treat 25 dry t/day to Class A biosolids in a stand alone VSD facility (similar to Concept 3 from Fig. 12.11).

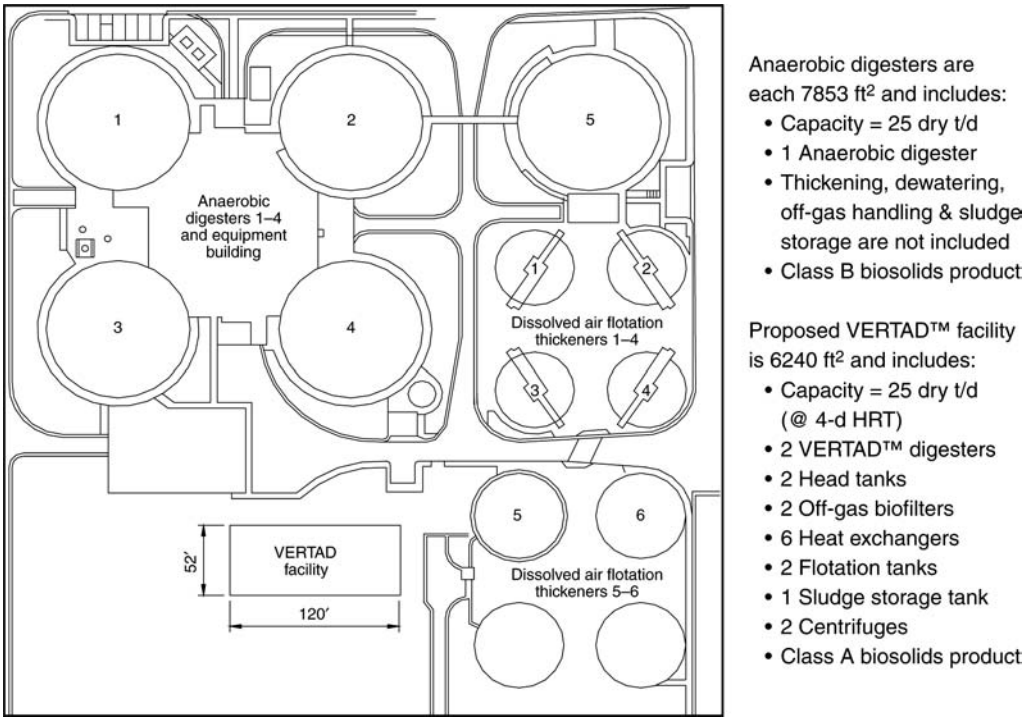


Fig. 12.10. Example of vertical shaft digestion retrofit at the king county south treatment plant.

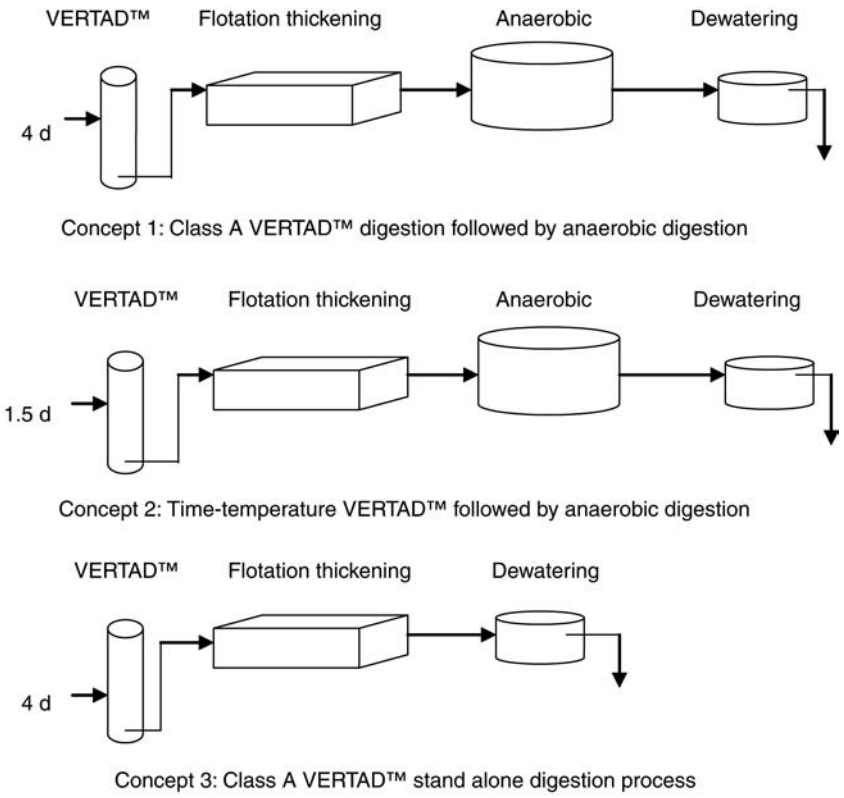


Fig. 12.11. Vertical shaft digestion (VERTAD™) process flow diagrams.

5.3.12. Full-Scale Design and Economics

The results of the demonstration project provided the basis for full-scale design parameters and cost estimates for the VSD process. Planning level designs were developed for three alternatives for solids treatment facilities at a planned future 36 MGD treatment plant in King County. The alternative flow sheets presented in Fig. 12.11 were developed in detail for the County (3, 47).

The present worth of capital costs for a system with VSD pre-treatment prior to anaerobic digestion was similar to that of mesophilic anaerobic digestion alone. The present worth of operating cost was significantly less than conventional anaerobic, primarily due to savings in dewatering and hauling cost in the VSD system.

Additional benefits not accounted for in the capital and operating cost analysis are expected to further improve the comparison, making VSD a favorable choice for King County or a similar community. These additional benefits include, but are not limited to:

1. The value of Class A biosolids (the potentially increased market for beneficial reuse)
2. Low grade heat recovered from the process can be utilized for space heating

3. Decreased land requirements for the VSD process
4. VSD product synergy with a subsequent anaerobic digestion step (improved mixing, less scum, higher solids concentration, decreased size of dewatering facility, improved dewatering)
5. Reduction in the NIMBY effect due to minimal odor release, and an aesthetically pleasing (out-of-sight) facility design

5.4. Capital Costs

Except for very small flow facilities, the capital cost of a VSD system is lower than that in conventional plants of similar size. Decreased land requirements, considerably less surface tankage (less concrete), less dewatering equipment and fewer pumps are some of the key elements decreasing the capital cost.

Several factors support the reduced capital costs and land requirements of VSD systems. These factors amount to VSD requiring 10% to 20% of the total land required for conventional anaerobic plants of equivalent capacity—reducing visual and environmental impact. Some of these factors include:

1. Eighty percent of the bioreactor volume is below grade—eliminating surface tankage.
2. Due to the high oxygen transfer efficiency in VSD systems, the residence time required in the bioreactor is decreased relative to conventional technologies—making the required reactor volume smaller.
3. The solids are easily float-thickened to 8% to 12% TS out of the VSD reactor. Float thickening in this manner significantly reduces the size of the downstream dewatering facility.

5.4.1. Operating Costs

The most significant savings realized in the VSD process relate to the aeration system (3). The basis of the VSD process is that the oxygen transfer efficiency is significantly higher than that in a conventional aerobic digestion system due to the pressure at the depth where air is introduced to the bioreactor. In a recent comparison study of the energy requirements between VSD and ATAD processes, it was found that VSD out-performed a conventional ATAD process, operating with 31% to 45% less energy per pound of VS destroyed in the system. It was also found that the VSD process obtained a doubling in oxygen transfer over the conventional ATAD system with 50% OTE compared to 24% OTE, respectively. These results are summarized in Table 12.9.

A VSD reactor operating at 4% solids can attain an oxygen transfer efficiency of approximately 50%. The resulting aeration power requirement is less than 1.3 kWh/kg (1200 kWh/t) of volatile solids destroyed or 0.35 kWh/kg (315 kWh/t) of total solids treated. No additional mixing energy is needed; therefore, the power requirement is much lower than the combined aeration and mixing power consumed by conventional aerobic processes.

This air that is economically and efficiently introduced to the bioreactor aids in several other process functions at no incremental cost. Not only does the air satisfy the primary requirement of providing the microbes with dissolved oxygen, it serves as an air lift pump—eliminating the need for mixers in the bioreactor. The air indirectly provides the dissolved gasses necessary for solids flotation in the flotation cell that follows the bioreactor—decreasing the size of the downstream dewatering equipment.

Table 12.9
Comparison of vertical shaft digestion and conventional autothermal thermophilic aerobic digestion

Parameter	ATAD (design)	ATAD (case study)	VERTAD™ (design)
Power usage (kWh/t TS fed)	442	520–641	315
Power usage (kWh/kg VS destroyed)	1.52	1.85–2.32	1.27
Aeration ($m^3/h/m^3$ active reactor volume)	4	Not measured	1.7
Time for VS destruction of 40% to 42% (day)	5–8	12–15	3.5–5
Average system OTE (%)	Not reported	24%	50%

Savings on operating costs have also been realized in the VSD system due to decreased chemical requirements. The VSD biosolids dewater to high cake solids with a very low polymer demand. VSD product can be dewatered to 30% to 35% solids using a conventional centrifuge, with less than 20 lb/t polymer addition. The exceptionally low polymer consumption reduces operating costs considerably. The ability to effectively dewater biosolids is extremely important due to the high costs associated with hauling and application or landfilling. The high solids content of the dewatered product reduces trucking and disposal costs thus again reducing operating costs. The nutrient value of the Class A biosolids product makes it favorable in any beneficial reuse program.

6. CONCLUSIONS

The following conclusions were made based on the results of the demonstration project (3):

1. The vertical shaft digestion (VSD commercially known as VERTAD™) reactor readily circulates thickened solids (4% to 6% TS); the upper zones are well mixed while the lower zone is hydraulically separate, providing strict adherence to the Class A pathogen requirements of US EPA 40 CFR 503.
2. The vector attraction reduction and pathogen destruction requirements for Class A biosolids were achieved with a 4 day solids retention time (US EPA 40 CFR 503, Alternative 1).
3. Oxygen transfer efficiency was greater than 50% when the reactor total solids concentration was at or below 4.5%.
4. VSD effluent could be easily float thickened to 8% to 12% TS by pH-shift CO_2 release; thickened product dewatered to greater than 30% cake solids with low polymer demand (14 b/t).
5. Organic nitrogen and fats, oils, and greases were preferentially degraded over organic solids comprised primarily of mesophilic anaerobic digestion of VSD effluent provided overall volatile solids destruction of 67% and gas production of 0.36 L CH_4/g COD removed with a combined

solids retention time of 15 days (4 day SRT in VSD followed by an 11 day SRT in anaerobic digestion).
6. VSD had low operating cost due to low energy requirements (1.27 kWh/kg VS destroyed), low polymer requirements (14 lb/t), and low trucking/disposal costs (≥30% TS cake).
7. A cost evaluation of full-scale implementation at King County treatment facilities indicated that a combined system of VSD and mesophilic anaerobic digestion has a similar present worth of capital and operating costs compared to traditional anaerobic digestion.
8. The VSD process has a minimal footprint requirement making it an ideal retrofit for facilities that require additional capacity, or current Class B biosolids generators that wish to produce Class A biosolids.

REFERENCES

1. J. Guild and D. Pollock, *Startup of a Full-Scale VERTREAT™ Wastewater Treatment Plant at a Cheese Production Facility*, NORAM Engineering and Constructors, Ltd., Vancouver, BC, Canada (2002).
2. J. Guild, D. Pollock, and D. Cameron, *Refinery Wastewater Treatment Utilizing the VERTREAT™ Process*, Technical Report, NORAM Engineering and Constructors, Ltd., Vancouver, BC, Canada, March (2003).
3. J. Guild, M. Boyle, L. Sasser, D. Pollock, and J. Yoo, VERTAD™—auto-thermophilic aerobic digestion: demonstration-scale test results, *Proceedings Water Environment Federation 74th Annual Conference & Exposition*, Atlanta, GA (2001).
4. N. K. Shammas and L. K. Wang, Aerobic digestion. In: *Biological Treatment Processes*, 2nd ed., L. K. Wang, N. C. Pereira, and Y. T. Hung (eds.), Humana Press, Totowa, NJ. pp. 635–669 (2009).
5. US EPA *Technology Assessment of the Deep Shaft Biological Reactor*, US Environmental Protection Agency, Washington, DC. Technical Report No. EPA/600/2-82-002 (1982).
6. P. G. Daly and C. C. Shen, The deep shaft biological treatment process, *Proceedings of Purdue Industrial Waste Conference*, p. 479, Purdue University, West Lafayette, IN (1989).
7. L. K. Wang and P. G. Daly, *Preliminary Design Report of a 10-MGD Deep Shaft-Flotation Plant for the City of Bango*r, US Department of Commerce, National Technical Information Service, Springfield, VA. Technical Reports Nos. PB88-200597/AS, 42 p.; and PB88-200605/AS, 171 p (1987).
8. N. K. Shammas, L. K. Wang, J. Guild, and D. Pollock, Vertical shaft bioreactors. In: *Advanced Biological Treatment Processes*, L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.), Humana Press, Totowa, NJ. pp. 59–108 (2009).
9. WEF *Advanced Digestion White Paper*, Advanced Digestion Technology Team (Chaired by Dave Parry), Bioenergy Subcommittee, Residuals and Biosolids Committee, Water Environment Federation, Alexandria, VA (2004).
10. L. K. Wang, L. Kurylko, and M. H. S. Wang, *Sequencing Batch Liquid Treatment*, US Patent No. 5354458, US Patent & Trademark Office, Washington, DC (1996).
11. L. K. Wang and Y. Li, Sequencing batch reactors, In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ. pp. 459–512 (2009).
12. L. K. Wang, E. Fahey, and Z. Wu, Dissolved air flotation, In: *Physicochemical Treatment Processes*, L. K. Wang, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ. pp. 431–500 (2005).

13. L. K. Wang and N. C. Pereira, *Biological Treatment Processes*, *Handbook of Environmental Engineering*, Vol. 3, Humana Press, Totowa, NJ. 498 p. (1986).
14. US EPA *Control of Air Emissions from Superfund Sites*, US Environmental Protection Agency, Washington, DC. Technical Report No. EPA/625/R-92/012 (1992).
15. L. K. Wang, N. C. Pereira, and Y. T. Hung (eds.), *Advanced Air and Noise Pollution Control*, Humana Press, Totowa, NJ. 526 p. (2005).
16. ACWPCP *Master Plan Update-Aerobic Digestion*, Arlington County Water Pollution Control Plant, Technical Memorandum XIII, April (2002).
17. US EPA *Sludge Handling and Conditioning*, US Environmental Protection Agency, Office of Water Program Operations, Washington, DC, EPA 430/9-78-002, February (1978).
18. D. A. Burke, Producing exceptional quality biosolids through digestion pasteurization and re-digestion, *Proceedings of WEF/AWWA/CWEA Joint Residuals and Biosolids Management Conference*, San Diego, CA (2001).
19. D. D. Drury et al., The use of three stage digestion to mitigate the adverse impacts of thermophilic digestion, *Proceedings of WEF/AWWA/CWEA Joint Residuals and Biosolids Management Conference*, San Diego, CA (2001).
20. G. L. Hernandez, et al., Hyperion advanced digestion pilot program, *Proceedings 16th Annual Residuals and Biosolids Management Conference*, Austin, TX (2002).
21. J. A. Oleszkiewicz and D. S. Mavinic, Wastewater biosolids: an overview of processing, treatment, and management, *Journal of Environmental Engineering Science*, 1, 75–88 (2002).
22. T. Valentine, In search of digester capacity for less cost? two-phase and three-phase digestion hold much promise, *Proceedings 16th Annual Residuals and Biosolids Management Conference*, Austin, TX (2002).
23. WEF, *Manual of Practice FD-9—Wastewater Residuals Stabilization*, Water Environment Federation, Alexandria, VA (1995).
24. US EPA *Autothermal Thermophilic Aerobic Digestion of Municipal Wastewater Sludge*. EPA-625/10-90-007, US Environmental Protection Agency, Cincinnati, OH (1990).
25. C. H. Burnett, Technology and process options for autothermal thermophilic aerobic digestion. *Proceedings Water Environment Federation 67th Annual Conference & Exposition*, Chicago, IL (1994).
26. S. Murthy, J. T. Novak, R.D. Holbrook, and F. Surovik, Mesophilic aeration of autothermal thermophilic aerobically digested biosolids to improve plant operations, *Water Environment Research*, 72 (2000).
27. R. Gemmell, R. Deshevy, M. Elliott, G. Crawford, and S. Murthy, Design considerations and operating experience for a full scale dual digestion system with separate waste activated sludge thickening, *Proceedings Water Environment Federation 73rd Annual Conference & Exposition*, Anaheim, CA (2000).
28. Metcalf & Eddy *Wastewater Engineering—Treatment, Disposal, and Reuse*, McGraw-Hill, New York (1991).
29. H. G. Kelly, H. Melcer, and D. S. Mavinic, Autothermal thermophilic aerobic digestion of municipal sludges: a one-year, full-scale demonstration project, *Water Environment Research*, 65, 849 (1993).
30. C. Beer, L. K. Wang, and L. J. Hetling, Full scale operations of plug flow activated sludge systems, *Journal of New England Water Pollution Control Association*, 9, 2, 145–173 (1975).
31. L. K. Wang and M. H. S. Wang, Control tests and kinetics of activated sludge process, *Water, Air and Soil Pollution*, 8, P315–351 (1977).

32. C. Beer, and L. K. Wang, Activated sludge systems using nitrate respiration—design considerations, *Journal of Water Pollution Control Federation*, 50, 9, 2120–2131 (1978).
33. C. Beer, L. J. Hetling, and L. K. Wang, *Full-Scale Operation of Plug Flow Activated Sludge Systems at Coxsackie WWTP*, Technical Report # 42, NY State Department of Environmental Conservation, Albany, NY (1975).
34. C. Beer and L. K. Wang, *Process Design of Single-Sludge Activated Sludge Systems Using Nitrate Respiration,* Technical Report No. 50 New York State Department of Environmental Conservation, Albany, NY *Proceedings of the 49th Annual Meeting of the New York Water Pollution Control Association*, NYC, January 17–19, Recipients of the Kenneth Research Award (1977).
35. C. Beer, J. F. Bergenthal, and L. K. Wang, A study of endogenous nitrate respiration of activated sludge, *Proceedings of the 9th Mid-Atlantic Industrial Waste Conference*, Bucknell University, Lewisburg, PA (1977).
36. L. K. Wang, Chemistry of nitrification-denitrification process, *Journal of Environmental Science*, 21, 23–28 (1978).
37. L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.), *Advanced Biological Treatment Processes*, Humana Press, Totowa, NJ. 738 p. (2009).
38. T. C. Rooney and N. A. Mignone, Influence of basin geometry on different generic types of aeration equipment, *Proceedings 33rd Purdue Industrial Waste Conference*, Ann Arbor Science, Ann Arbor, MI (1978).
39. Y. Shang and S. Shihwu, Thermophilic-mesophilic two-stage anaerobic digestion system application, *Proceedings of WEF/AWWA/CWEA Joint Residuals and Biosolids Management Conference*, San Diego, CA (2001).
40. G. Winfield, Design of an acid/gas digestion process for Baltimore's Back River plant, *Proceedings of WEF/AWWA/CWEA Joint Residuals and Biosolids Management Conference*, WEF, San Diego, CA (2001).
41. J. R. Taricska, P. Chen, Y. T. Hung, and D. Long, Anaerobic digestion, In: *Biological Treatment Processes*, 2nd ed., L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ. p. 589–634 (2009).
42. US EPA *Innovative and Alternative Technology Assessment Manual*, US Environmental Protection Agency, Washington, DC, Technical Report No. EPA430/9-78-009 (1980).
43. L. K. Wang, C. Williford, W. Y. Chen, N. K. Shammas, and G. P. Sakellaropoulos, Evaporation processes, In: *Advanced Physicochemical Treatment Processes*, L. K. Wang, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ. pp. 549–580 (2006).
44. L. K. Wang, N. C. Pereira, and Y. T. Hung (eds.), *Air Pollution Control Engineering*, Humana Press, Totowa, NJ. 504 p. (2004).
45. G. Leson and A. Winter, Biofiltration: an innovative air pollution control technology for VOC emissions, *JAWMA*, 41, 8 (1991).
46. US ACE, *Civil Works Construction Cost Index System Manual*, US Army Corps of Engineers, Washington, DC, PP 44 (2008-Tables) (2008).
47. E&A Environmental Consultants *VERTAD™ Demonstration Project-Aerobic Thermophilic Digestion in a Deep Vertical Reactor*, Final Evaluation Report Prepared for King County Department of Natural Resources, Project 30900, Seattle, WA, May 25 (2001).
48. Dames and Moore *Geotechnical Survey Pilot Scale Wastewater Solids Processing Project*, Dames and Moore, New Jersey (1994).
49. D. A. Burke, R. Butle, and S. Hummel, An assessment of the AGF (Anoxic Gas Flotation) High Rate Anaerobic Digestion Process, *Proceedings Water Environment Federation 12th Annual Residuals and Biosolids Management Conference*, Bellevue, WA (1998).

APPENDIX

United States Yearly Average Cost Index for Utilities US Army Corps of Engineers (46)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	538.74
1987	353.35	2008	552.16

13
Land Application of Biosolids

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract Biosolids are essentially organic materials produced during wastewater treatment which may be put to beneficial use. A popular example of such use is the addition of biosolids to soil to supply nutrients and replenish soil organic matter. Biosolids can be applied on agricultural land, forests, rangelands, or on disturbed land in need of reclamation. Recycling biosolids through land application serves several purposes. It improves soil properties, such as texture and water holding capacity, which make conditions more favorable for root growth and increases the drought tolerance of vegetation. Biosolids application also supplies nutrients

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_13

essential for plant growth, including nitrogen and phosphorous, as well as some essential micro nutrients such as nickel, zinc, and copper.

In addition to describing the methods for land application of biosolids, the chapter covers its advantages and disadvantages, design criteria, performance, costs of recycling through land application, biosolids disposal on land (landfill), biosolids landfill methods, preliminary planning, facility design, operation and maintenance, site closure, costs of biosolids disposal on land (landfill) and application examples.

Key Words Beneficial uses • biosolids • design and costs • land application • landfill • planning.

1. INTRODUCTION

Biosolids are essentially organic materials produced during wastewater treatment which may be put to beneficial use. A popular example of such use is the addition of biosolids to soil to supply nutrients and replenish soil organic matter. Biosolids can be applied on agricultural land, forests, rangelands, or on disturbed land in need of reclamation (1). The thrust of recent legislation has been to encourage such beneficial recycling of biosolids through land application (2). The establishment of the industrial waste pretreatment programs (3) with the objective of reducing toxic pollutant loadings to municipal treatment facilities rendered more municipal biosolids suitable for reuse.

Wastewater biosolids may not always be used as a resource because of land acquisition constrains, the unavailability of agricultural nearby land or because they contain high levels of metals and other toxic substances. Since ocean disposal is no longer considered a viable or an appropriate alternative to utilization, land disposal through landfill has been optimized so that the increasing amounts of biosolids generated by wastewater treatment plants can be accepted. Development of formalized methods for biosolids disposal to land is recent. Major efforts in this area have been funded by the US EPA since 1974 (4).

US EPA regulations (5) under Title 40 Code of Federal Regulations Part 503 (40 CFR 503) and their amendments (6), established the minimum national standards for the use and disposal of municipal biosolids. The reader is referred to Refs. (7–11) for discussion and detailed information on the background, guidance, risk assessment and applications of the regulations to control the recycling of biosolids through land application and disposal in landfills.

2. RECYCLING OF BIOSOLIDS THROUGH LAND APPLICATION

Recycling biosolids through land application serves several purposes. It improves soil properties, such as texture and water holding capacity, which make conditions more favorable for root growth and increases the drought tolerance of vegetation. Biosolids application also supplies nutrients essential for plant growth, including nitrogen and phosphorous, as well as some essential micro nutrients such as nickel, zinc, and copper (2). Biosolids can also serve as an alternative or substitute for expensive chemical fertilizers. The nutrients in the biosolids offer several advantages over those in inorganic fertilizers because they are organic and are released slowly to growing plants. These organic forms of nutrients are less water soluble and, therefore, less likely to leach into groundwater or run off into surface waters (1).

Land application is well-suited for managing solids from any size wastewater treatment facility. As the method of choice for small facilities, it offers cost advantages, benefits to the environment, and value to the agricultural community. However, biosolids produced by many major metropolitan areas across the country are also land applied. For example, biosolids from the Blue Plains Wastewater Treatment Facility serving the District of Columbia and surrounding communities in Virginia and Maryland have been land applied since the plant began operation in 1930. The cities of Philadelphia, Chicago, Denver, New York, Seattle, and Los Angeles all land apply at least part of their biosolids production (1).

Land application is most easily implemented where agricultural land is available near the site of biosolids production, but advances in transportation have made land application viable even where hauling distances are greater than 1000 miles. For example, Philadelphia hauls dewatered biosolids 250 miles to reclaim strip-mines in western Pennsylvania and New York City ships some of its biosolids over 2000 miles to Texas and Colorado (1).

3. DESCRIPTION

There are several methods for land applications of biosolids. The selection of the method depends on the type of land and the consistency of the biosolids. Liquid biosolids are essentially 94% to 97% water with relatively low amounts of solids (3% to 6%). These can be injected into the soil or applied to the land surface. Specialized vehicles or modified tanker trucks are used to inject biosolids into the soil. These tankers have hoses leading from the storage tank to injection nozzles which release the biosolids. Biosolids applied to the land surface are usually incorporated into the soil with conventional farm equipment.

It is often economical to reduce the volume of biosolids prior to transportation or storage. The amount of water in biosolids can be reduced through mechanical processes such as draining, pressing, or centrifuging, resulting in a material composed of up to 30% dry solids (12–15). This material will be the consistency of damp soil. Dewatered biosolids do not require any specialized equipment and can be applied with conventional agricultural equipment, such as manure spreaders pulled by tractors.

The Environmental Protection Agency's 40 CFR Part 503, *Standards for the Use and Disposal of Sewage Sludge* (5), requires that wastewater solids be processed before they are land applied. This processing is referred to as stabilization and helps minimize odor generation, destroys pathogens (disease causing organisms) and reduces vector attraction potential. There are several methods to stabilize wastewater solids, including (12, 16):

(a) Adjustment of pH, lime, or alkaline stabilization
(b) Anaerobic digestion
(c) Aerobic digestion
(d) Composting
(e) Heat drying

The Part 503 Rule (5) defines two types of biosolids with respect to pathogen reduction, Class A and Class B, depending on the degree of treatment the solids have received. Both types are safe for land application, but additional requirements are imposed on Class B materials. These are detailed in the Part 503 Rule and include such things as restricting public access

Table 13.1
Maximum metal concentrations

Metal	Ceiling concentration (mg/kg)	Cumulative pollutant loading rates (kg/ha)	Pollutant concentrations (mg/kg)
Arsenic	75	41	41
Cadmium	85	39	39
Copper	4300	1500	1500
Lead	840	300	300
Mercury	57	17	17
Molybdenum	75	*NL*	*NL*
Nickel	420	420	420
Selenium	100	100	100
Zinc	7500	2800	2800

NL = No limit.
Source: US EPA.

to the application site, limiting livestock grazing, and controlling crop harvesting schedules. Class A biosolids (biosolids treated so that there are no detectable pathogens) are not subject to these restrictions.

In addition to stabilization, the Part 503 Rule sets maximum concentrations of metals which cannot be exceeded in biosolids that will be land applied. These are termed Ceiling Concentrations. Part 503 also establishes Cumulative Pollutant Loading Rates for eight metals which may not be exceeded at land application sites. A third set of metals criteria is also included in Part 503, known as Pollutant Concentrations. If these concentrations are not exceeded in the biosolids to be land applied, the Cumulative Pollutant Loading Rates do not need to be tracked. Table 13.1 shows the three sets of federal limits applicable to biosolids to be land applied (5, 17).

The term *Exceptional Quality* is often used to describe a biosolids product which meets Class A pathogen reduction requirements, the most stringent metals limits (Pollutant Concentrations), and vector attraction reduction standards specified in the Part 503 Rule. Vectors (flies, mosquitoes, rodents, birds, etc.) can transmit diseases directly to humans or play a specific role in the life cycle of a pathogen as a host. Vector attraction reduction refers to processing which makes the biosolids less attractive to vectors thereby reducing the potential for transmitting diseases. Exceptional Quality biosolids products are as safe as other agricultural and horticultural products and may be used without site restrictions.

4. ADVANTAGES AND DISADVANTAGES

Land application offers several advantages as well as some disadvantages that must be considered before selecting this option for managing biosolids (1).

Land application is an excellent way to recycle wastewater solids as long as the material is quality-controlled. It returns valuable nutrients to the soil and enhances conditions for vegetative growth. Land application is a relatively inexpensive option and capital investments are generally lower than other biosolids management technologies. Contractors can provide

the necessary hauling and land application equipment. In addition, on-site spatial needs can be relatively minor depending on the method of stabilization selected.

Although land application requires relatively less capital, the process can be labor intensive. Even if contractors are used for application, management oversight is essential for program success. Land application is also limited to certain times of the year, especially in colder climates. Biosolids should not be applied to frozen or snow covered grounds, while farm fields are sometimes not accessible during the growing season. Therefore, it is often necessary to provide a storage capacity in conjunction with land application programs. Even when the timing is right (for example, prior to crop planting in agricultural applications) weather can interfere with the application. Spring rains can make it impossible to get application equipment into farm fields, making it necessary to store biosolids until weather conditions improve.

Another disadvantage of land application is potential public opposition, which is encountered most often when the beneficial use site is close to residential areas. One of the primary reasons for public concern is odor. In worst case situations, municipalities, or counties may pass ordinances which ban or restrict the use of biosolids. However, many successful programs have gained public support through effective communications, an absolutely essential component in the beneficial use of biosolids.

Despite many positive impacts to the environment, land application can have negative impacts on water, soil, and air if not practiced correctly.

Negative impacts to water result from the application of biosolids at rates that exceed the nutrient requirements of the vegetation. Excess nutrients in the biosolids (primarily nitrogen compounds) can leach from the soil and reach groundwater. Runoff from rainfall may also carry excess nutrients to surface water. However, because biosolids are a slow release fertilizer, the potential for nitrogen compounds to leach from biosolids amended soil is less than that posed by the use of chemical fertilizers. In areas fertilized by either biosolids or chemicals, these potential impacts are mitigated by proper management practices, including the application of biosolids at agronomic rates (the rate nutrients are used by the vegetation.) Maintenance of buffer zones between application areas and surface water bodies and soil conservation practices will minimize impacts to surface water.

Negative impacts to soil can result from mismanagement of a biosolids land application. Federal regulations contain standards related to all metals of concern and application of biosolids which meets these standards should not result in the accumulation of metals to harmful levels. Stringent record keeping and reporting requirements on both the federal and state level are imposed to prevent mismanagement.

Odors from biosolids applications are the primary negative impact to the air. Most odors associated with land application are a greater nuisance than threat to human health or the environment. Odor controls focus on reducing the odor potential of the biosolids or incorporating them into the soil. Stabilization processes such as digestion can decrease the potential for odor generation. Biosolids that have been disinfected through the addition of lime may emit ammonia odors but they are generally localized and dissipate rapidly. Biosolids stabilization reduces odors and usually results in an operation that is less offensive than manure application.

Overall, a properly managed biosolids land application program is preferable to the use of conventional fertilizers for the following reasons (1):

(a) Biosolids are a recycled product, use of which does not deplete non-renewable resources such as phosphorous.
(b) The nutrients in biosolids are not as soluble as those in chemical fertilizers and are therefore released more slowly.
(c) Biosolids appliers are required to maintain setbacks from water resources and are often subject to more stringent soil conservation and erosion control practices, nutrient management, and record keeping and reporting requirements than farmers who use only chemical fertilizers or manures.
(d) Biosolids are closely monitored.
(e) The organic matter in biosolids improves soil properties for optimum plant growth, including tilth, friability, fertility, and water holding capacity. They also decrease the need for pesticide use.

A joint policy statement of the US Department of Agriculture, the US Food & Drug Administration, and the US Environmental Protection Agency states, "...the use of high quality biosolids coupled with proper management procedures, should safeguard the consumer from contaminated crops and minimize any potential adverse effect on the environment" (18).

5. DESIGN CRITERIA

Design criteria for land application programs address issues related to application rates and suitable sites. Design criteria for physical facilities (such as stabilization) that are part of land application programs are discussed in other chapters. Biosolids, site, and vegetative characteristics are the most important design factors to consider.

Biosolids must meet regulatory requirements for stabilization and metals content. In addition, nutrient content and physical characteristics, such as percent solids, are used to determine the appropriate application rate for the crop that will be grown and the soil in which the crops will be grown.

Site suitability is determined based on such factors as soil characteristics, slope, depth to groundwater, and proximity to surface water. In addition, many states have established site requirements to further protect water quality. Some examples include:

(a) Sufficient land to provide areas of non-application (buffers) around surface water bodies, wells, and wetlands
(b) Depth from the soil surface to groundwater equal to at least 1 m
(c) Soil pH in the range of 5.5 to 7.5 to minimize metal leaching and maximize crop growing conditions

Site suitability is also influenced by the character of the surrounding area. While odors and truck traffic many not be objectionable in an agricultural area, both will adversely impact residential developments and community centers close to fields where biosolids are applied.

The type of vegetation to be grown is also a design consideration. Vegetation, like soil characteristics, will generally not exclude biosolids application since most vegetation will benefit from the practice. However, the type of vegetation will impact the choice of application equipment, the amount of biosolids to be applied, and the timing of applications. The amount

Table 13.2
Typical biosolids application scenarios

Type of site/ vegetation	Schedule	Application frequency	Application rate
Agricultural land			
Corn	April, May, after harvest	Annually	5 to 10 dry tons per acre
Small grains	March-June, August, fall	Up to 3 times per year	2 to 5 dry tons per acre
Soybeans	April-June, fall	Annually	5 to 20 dry tons per acre
Hay	After each cutting	Up to 3 times per year	2 to 5 dry tons per acre
Forest land	Year round	Once every 2 to 5 years	5 to 100 dry tons per acre
Range land	Year round	Once every 1 to 2 years	2 to 60 dry tons per acre
Reclamation sites	Year round	Once	60 to 100 dry tons per acre

Source: US EPA.

of biosolids that may be applied to a site is a function of the amount of nutrients required by the vegetation and the amount of metals found in the biosolids. Table 13.2 summarizes the application frequency, timing, and rates for various types of sites (1, 17).

Another factor to be considered in designing a land application program is the timing of applications. Long periods of saturated or frozen ground limit opportunities for application. This is an important consideration in programs using agricultural lands; applications must be performed at times convenient to the farmer and must not interfere with the planting of crops. Most application of biosolids to agricultural land occurs in the early spring or late fall. As a result, storage or an alternate biosolids management option must be available to handle biosolids when application is not possible. Forest lands and reclamation sites allow more leeway in the timing of applications. In some areas of the United States, application can proceed year round.

Application is most beneficial on agricultural land in late fall or early spring before the crop is planted. Timing is less critical in forest applications when nutrients can be incorporated into the soil throughout the growing period. Winter application is less desirable in many locales. Rangelands and pasturelands also are more adaptable to applications during various seasons. Applications can be made as long as ground is not saturated or snow covered and whenever livestock can be grazed on alternate lands for at least 30 days after the application. The timing of single applications in land reclamation programs is less critical and may be dictated by factors such as regulatory compliance schedules.

6. PERFORMANCE

In 1995, approximately 54% of wastewater treatment plants managed biosolids through land application, an increase of almost 20% from information reported in 1993 (5, 19). The vast majority of these land application programs use agricultural land, with minor amounts applied to forest lands, rangelands, or land in need of reclamation.

The use of land application increased steadily in the 1980s for several reasons, including decreasing availability and increasing costs associated with landfill disposal. Research

also helped refine procedures for proper land application. Meanwhile, implementation of the Nationwide Pretreatment Program (3) resulted in significant improvements in biosolids quality. The 1993 adoption of the Part 503 Rule created a structure for consistent application procedures across the nation. The regulations were developed with input from the US Department of Agriculture, the US Food and Drug Administration, biosolids generators, environmental groups, the public, state regulators, and academic researchers. Conservative assumptions were used to create regulations to "protect public health and the environment from all reasonably anticipated adverse effects" (5).

Land application is a reliable biosolids management option as long as the system is designed to address such issues as storage or alternate management for biosolids during periods when application cannot take place due to unfavorable weather or field conditions. Public opposition rather than technical constraints is the most common reason for discontinuing land application programs (1). Martha Prothro, a Former Deputy Assistant Administrator for Water, US Environmental Protection Agency stated that (20) "*In fact, in all the years that properly treated biosolids have been applied to the land, we have been unable to find one documented case of illness or disease that resulted.*"

Land application systems generally use uncomplicated, reliable equipment. Operations include pathogen reduction processing, dewatering, loading of transport vehicles, transfer to application equipment, and the actual application. Operations and maintenance considerations associated with pathogen reduction processing are discussed in Refs. (12, 16). The other operations require labor skills of heavy equipment operators, equipment maintenance personnel, and field technicians for sampling, all normally associated with wastewater treatment facilities.

In addition, the biosolids generator is responsible for complying with state and local requirements as well as federal regulations. The biosolids manager must be able to calculate agronomic rates and comply with record keeping and recording requirements. In fact, the generator and land applier must sign certification statements verifying accuracy and compliance (1). The generator should also allocate time to communicate with farmers, landowners, and neighbors about the benefits of biosolids recycling. Control of odors, along with a viable monitoring program, is most important for public acceptance (21).

Detailed discussions and more information on biosolids recycling for land application can be found in Refs. (22–27)

7. COSTS OF RECYCLING THROUGH LAND APPLICATION

It is difficult to estimate the cost of land application of biosolids without specific program details. For example, there is some economy of scale due to large equipment purchases. The same size machine might be needed for a program that manages 10 t/day of dry biosolids as one managing 50 dry t/day; the cost of that machine can be spread over the 10 or 50 dry tons, greatly affecting average costs per dry ton. One source identified costs for land application varying from USD 60 to USD 290/dry ton in 1996 (28), which is equivalent to 2008 USD 75 to USD 360 (Appendix, 29). This range reflects the wide variety in land application methods as well as varying methods to prepare biosolids for land application. For example,

costs for programs using dewatered biosolids include an additional step whereas costs for programs using liquid biosolids do not reflect the cost of dewatering. They do, however, include generally higher transportation costs.

Despite the wide range of costs for land application programs, several elements must be considered in estimating the cost of any biosolids land application program (1):

(a) Purchase of application equipment or contracting for application services
(b) Transportation
(c) Equipment maintenance and fuel
(d) Loading facilities
(e) Labor
(f) Capital, operation and maintenance of stabilization facilities
(g) Ability to manage and control odors
(h) Dewatering (optional)
(i) Storage or alternate management option for periods when application is not possible due to weather or climate
(j) Regulatory compliance, such as permit applications, site monitoring, and biosolids analyses
(k) Public education and outreach efforts

Land must also be secured. Some municipalities have purchased farms for land application; others apply biosolids to privately held land. Some operating costs can be offset through the sale of the biosolids material. Since the biosolids reduce the need for fertilizers and pH adjustment, farmers pay to have biosolids applied to their lands.

8. BIOSOLIDS DISPOSAL ON LAND (LANDFILL)

Biosolids landfill can be defined as the planned burial of wastewater solids and processed biosolids at a designated land site. The solids are placed into a prepared site or excavated trench and covered with a layer of soil. The soil cover must be deeper than the depth of the plow zone (about 8 to 10 in.). For the most part, landfilling of screenings, grit, and ash is accomplished with methods similar to those used for biosolids landfilling (4).

9. BIOSOLIDS LANDFILL METHODS

Biosolids landfill methods can be grouped into three general categories:

(a) Biosolids-only trench fill
(b) Biosolids-only area fill
(c) Co-disposal with refuse

General site and design criteria are discussed under these categories. A detailed discussion of biosolids landfills can be found in the US EPA Technology Transfer Process Design Manual, Municipal Sludge Landfills (30) and in Office of Solid Waste Report, Disposal of Sewage Sludge into a Sanitary Landfill (31).

9.1. Biosolids-Only Trench Fill

Stabilized or unstabilized biosolids are placed within a subsurface excavation and covered with soil. Trench operations are more specifically categorized as follows: narrow trench and

Table 13.3
Comparison of design criteria for narrow and wide trench landfill

Design criteria	Narrow trench (less than 10 ft)	Wide Trench (more than 10 ft)
Sludge solids content	15% to 20% for 2 to 3 ft widths, 20% to 28% for 3 to 10 ft widths	20% to 28% for land-based equipment; more than 28% for sludge based equipment
Ground slopes	Less than 20%	Less than 10%
Cover soil thickness	2 to 3 ft for 2 to 3 ft widths; 3 to 4 ft for 3 to 10 ft widths	3 to 4 ft for land based equipment; 4 to 5 ft for sludge based equipment
Sludge application rate	1200 to 5600 yd^3/acre	3200 to 14, 500 yd^3/acre
Equipment	Backhoe with loader, excavator, trenching machine	Track loader, dragline, scraper, track dozer

Source: US EPA.

wide trench. Narrow trenches are defined as having widths less than 10 ft; wide trenches are defined as having widths greater than 10 ft. The width of the trench is determined by the solids content of the receiving biosolids and its capability of supporting cover material and equipment. Distances between trenches should be large enough to provide sidewall stability, as well as space for soil stockpiles, operating equipment and haul vehicles.

Design considerations should include provisions to control leachate and gas migration, dust, vectors, and/or aesthetics. Leachate control measures include the maintenance of 2 to 5 ft of soil thickness between trench bottom and highest groundwater level or bedrock (2 ft for clay to 5 ft for sand), or membrane liners and leachate collection and treatment system. Installation of gas control facilities may be necessary if inhabited structures are nearby. A comparison of the three types of trench fill (32) is shown in Table 13.3.

9.1.1. Narrow Trenches

Trenches are defined as narrow when their widths are less than 10 ft (3 m). Biosolids are disposed in a single application and a single layer of cover soil is applied on top. Trenches are usually excavated by equipment based on solid ground adjacent to the trench, and equipment does not enter the excavation. Backhoes, excavators, and trenching machines are particularly useful. Excavated material is usually immediately applied as cover over an adjacent biosolids-filled trench. Biosolids are placed in trenches either directly from haul vehicles, through a chute extension, or by pumping. The main advantage of a 2 to 3 ft narrow trench is its ability to handle biosolids with relatively low solids content (15% to 20%). Instead of sinking to the bottom of the biosolids, the cover soil bridges over the trench and receives support from undisturbed soils along each side of the trench. A 3 to 10 ft width is more appropriate for biosolids with solids content of 20% to 28%, which is high enough to support cover soil.

The application rates range from 1200 to 5600 yd^3 of biosolids/acre (2270 to 10,580 m^3/ha). Excavated material can be either used immediately to cover an adjacent biosolids—filled trench or stockpiled alongside and used to cover the trench from which it was removed. The surface soil cover thickness is about 4 ft (1.3 m).

9.1.2. Wide Trenches

Trenches are defined as wide when they have widths greater than 10 ft (3 m). Trenches are usually excavated by equipment operating inside the trench. Track loaders, draglines, scrapers, and track dozers are suitable. Excavated material is stockpiled on solid ground adjacent to the trench for subsequent application as cover material. If biosolids are incapable of supporting equipment, cover is applied by equipment based on solid undisturbed ground adjacent to the trench. A front-end loader is suitable for trenches up to 10 ft wide; a dragline is suitable for trench widths up to 50 ft. If biosolids can support equipment, a track dozer applies cover from within the trench.

Biosolids are placed in trenches by one of the two methods; from haul vehicles directly entering the trench and haul vehicles dumping from the top of the trench. Dikes can be used to confine biosolids to a specific area in a continuous trench. Disposal in wide trenches is suitable for biosolids with solids content of 20% or greater. The application rates range from 3200 to 14,500 yd^3 of biosolids/acre (6050 to 27,400 m^3/ha).

The surface cover thickness depends on the solids concentration of the biosolids. The covered biosolids will only be capable of supporting equipment when the solids concentration of the biosolids exceeds 25% to 30% and the biosolids have been topped with 3 to 5 ft (1 to 2 m) of soil.

The wide trench method has two distinct advantages; it is less land-intensive than the narrow trench method and groundwater protection can be provided by liners. The use of liners permits deeper excavations. The primary disadvantage of the wide trench method is the need for biosolids concentrations of greater than 20% solids. Biosolids with solid contents of greater than 30% to 35% will not flow, and extra effort is therefore required to spread them evenly in the trench. After maximum settlement has occurred in approximately 1 year, the area should be regraded to ensure proper drainage.

9.2. Biosolids-Only Area Fill

In the biosolids-only area fill method, the biosolids are mixed with soil and the mixture is placed on the original ground surface. This method requires substantial amounts of imported soil but may be suitable in areas where groundwater is shallow (liners can be easily installed) or bedrock prevails (that is, where excavation is neither possible nor required). Stabilized biosolids are best suited for this method, since daily cover is not usually provided.

To achieve stability and soil bearing capacity, sludge is mixed with a bulking agent, usually soil. The soil absorbs excess moisture from the sludge and increases its workability. The large quantities of soil required may require hauling from elsewhere. Provisions must be made to keep the stockpiled soil dry. Installation of a liner is generally required for groundwater control. Provisions are made for surface drainage control to prevent contamination of nearby surface waters, gas migration, dust, vectors and/or aesthetics. A comparison of the three types of area fill (32) is shown in Table 13.4.

9.2.1. Area Fill Mound

Area fill mound applications are generally suitable for stabilized biosolids with solids concentrations of 20% or more. Biosolids are mixed with a bulking agent, usually soil, and

Table 13.4
Comparison of design criteria for area fill mound, area fill layer and diked containment landfill

Design criteria	Area fill mound	Area fill layer	Diked containment
Sludge solids content	Greater than 20%	Greater than 15%	20% to 28% for land-based equipment; more than 28% for sludge-based equipment.
Sludge characteristics	Stabilized	Stabilized	Stabilized or unstabilized.
Ground slopes	No limitation if suitably prepared	Level ground preferred	Level ground or steep terrain if suitably prepared
Bulking required	Yes	Yes	Occasionally
Bulking ratio soil: sludge	0.5 to 2 soil:1 sludge	0.25 to 1 soil: 1 sludge	0 to 0.5 soil:1 sludge
Sludge application rate	3000 to $14,000\ yd^3$/acre	2000 to $9000\ yd^3$/acre	4800 to $15,000\ yd^3$/acre
Equipment	Track loader, backhoe with loader, track dozer	Track dozer, grader, track loader	Dragline, track dozer, soraper

Source: US EPA.

the mixture is hauled to the filling area, where it is stacked in mounds approximately 6 ft high. Cover material is then applied in a 3 ft thickness. This cover thickness may be increased to 5 ft if additional mounds are applied atop the first lift. The appropriate sludge/soil bulking ratio and soil cover thickness depend upon the solids content of the sludge as received, the need for mound stability and bearing capacity as dictated by the number of lifts and equipment weight. Lightweight equipment with swamp pad tracks is appropriate for area fill mound operations; heavier wheel equipment is appropriate in transporting bulking material to and from stockpiles. A level area is required for disposal; however, the use of earthen containment structures permits disposal in hilly areas.

9.2.2. Area Fill Layer

Area fill layer applications are suitable for stabilized biosolids with solids as low as 15%. Soil is mixed with biosolids, either at the filling area or at a special mixing area. The biosolids/soil mixture is spread in even layers of approximately 1 ft (0.3 m) thick, and 3 to 5 ft (1 to 1.5 m) of soil are added for final cover. Lightweight equipment with swamp pad tracks is appropriate for area fill layer operations; heavier wheel equipment is appropriate for hauling soil. Slopes should be relatively flat to prevent sludge from flowing downhill. However, if sludge solids content is high and/or sufficient bulking soil is used, the effect can be prevented and layering performed on mildly sloping terrain.

9.2.3. Dike Containment

Dike containment applications require biosolids with solids content of 20% or greater. This method is suitable for either stabilized or unstabilized biosolids. If the disposal site is level, earthen dikes are used on all four sides of the containment area. If the site is at the toe of the hill, only a partial diking is required. Access is provided to the top of the dike so that haul vehicles can dump biosolids directly into the containment. Depending on the type of equipment used, the interim cover will vary from 1 to 3 ft (0.3 to 1.0 m) and the final cover from 3 to 5 ft (1.0 to 1.5 m).

Cover material is applied either by a dragline based on solid ground atop the dikes or by track dozers directly on top of the sludge, depending upon sludge bearing capacity. Usually, operations are conducted without the addition of soil bulking agents, but occasionally soil bulking is added. Typical dimensions: 50 to 100 ft wide, 100 to 200 ft long, 10 to 30 ft deep. Although diked containment is an efficient disposal method from the standpoint of land use, it may necessitate controls for leachate outbreaks.

9.3. Co-Disposal with Refuse

The term co-disposal is used when municipal biosolids are disposed of at a refuse landfill. There are distinct trade-offs in using co-disposal method rather than the biosolids-only methods.

Biosolids can be disposed of in this manner if they are mixed with refuse or with soil. Mixing techniques are discussed in detail in the US EPA Office of Solid Waste Report, Disposal of Wastewater Biosolids into a Sanitary Landfill (31).

9.3.1. Biosolids/Refuse Mixture

Stabilized or unstabilized biosolids with solids content of 3% or greater are mixed with the refuse. Normally biosolids content is approximately 10% of the biosolids/refuse mixture. The biosolids are applied on top of the refuse at the working face of the landfill. The biosolids and refuse are thoroughly mixed before they are spread, compacted, and covered with soil. An interim cover of approximately 1 ft (0.3 m) and a final cover of 2 ft (0.6 m) is used. Application rates range from 500 to 4200 yd^3 of biosolids/acre (950 to 7900 m^3/ha).

9.3.2. Biosolids/Soil Mixture

In this operation, biosolids are mixed with soil and the mixture is used as cover for a refuse landfill. This method requires stabilized biosolids with at least 20% solids content. It promotes vegetation growth over completed landfill areas without the use of fertilizer. However, it may cause odors, since the biosolids are not completely buried. A final soil cover could be added if necessary to eliminate this problem.

Some wastewater treatment biosolids may not be suitable for landfilling by any of the methods described above. For biosolids-only landfills, the solids concentration should be 15% or more. Although soil may be used as a bulking agent to effectively increase the solids concentration to this level, cost-effectiveness may become a problem. Solids concentrations down to 3% are tolerated for co-disposal, but the absorptive capacity of the refuse should not be exceeded. An assessment of the suitability of various biosolids types is given in Table 13.5. In general, only stabilized and dewatered biosolids are recommended for landfill disposal.

Table 13.5
Suitability of biosolids for landfill

	Sludge only landfilling		Co-disposal landfilling	
Type of sludge	Suitability	Reason	Suitability	Reason
Liquid - unstabilized				
Gravity thickened primary, WAS and primary, and WAS	NS	OD, OP	NS	OD, OP
Flotation thickened primary and WAS, and WAS without chemicals	NS	OD, OP	NS	OD, OP
Flotation thickened WAS with chemicals	NS	OP	NS	OD, OP
Thermal conditioned primary or WAS	NS	OD, OP	MS	OD, OP
Liquid - stabilized				
Thickened anaerobic digested primary and primary, and WAS	NS	OP	MS	OP
Thickened aerobic digested primary and primary, and WAS	NS	OP	MS	OP
Thickened lime stabilized primary and primary, and WAS	NS	OP	MS	OP
Dewatered - unstabilized				
Vacuum filtered, lime conditioned primary	S	-	S	-
Dewatered - stabilized				
Drying bed digested and lime stabilized	S	-	S	-
Vacuum filtered, lime conditioned digested	S	-	S	-
Pressure filtered, lime conditioned digested	S	-	S	-
Centrifuged, digested and lime conditioned digested	S	-	S	-
Heat dried				
Heat dried digested	S	-	S	-
High temperature processed				
Incinerated dewatered primary and primary, and WAS	S	-	S	-
Wet-air oxidized primary and primary, and WAS	NS	OD, OP	MS	OD, OP

WAS - Waste-activated sludge.
NS - Not suitable.
MS - Marginally suitable.
S - Suitable.
OD - Odor problems.
OP - Operational problems.
Source: US. EPA.

9.4. Landfilling of Screenings, Grit, and Ash

Screenings and grit normally contain some putrescible materials and should be covered every day. Odors from temporarily uncovered solids may be alleviated by sprinkling the solids with lime. Special care should be exercised to assure vector control (e.g., safe poisons for rodent control, spraying for flies, and animal-proof fencing to keep pets from the area).

Residues (ash from the combustion of municipal wastewater solids) generally contain high concentrations of trace metals. Leachate from sites where incinerator ash is landfilled must be controlled to prevent metals contamination of groundwater.

10. PRELIMINARY PLANNING

The purpose of the preliminary planning activity is to select a disposal site and suitable method(s) of disposal. Preliminary planning is followed by detailed design, initial site development, site operation and maintenance, and final site closure.

Site selection is the major activity during the preliminary planning phase. Since the selection of a site is not completely independent of the selection of a method, the preliminary planning phase should also include the determination of biosolids characteristics and the identification of alternate landfill methods for each site.

10.1. Biosolids Characterization

Biosolids must be characterized as to quantity and quality. An estimate of the average biosolids quantity is necessary to establish landfill area requirements and the probable life of the disposal site. Data on minimum and maximum biosolids quantities are important for developing an understanding of daily operating requirements. Maximum daily biosolids quantities will govern equipment and storage facility sizing and daily operating schedules.

The character of the biosolids to be landfilled is directly related to the choice of a landfill method. Biosolids quality and the corresponding leachate can be roughly correlated; design of leachate treatment facilities is more effective if biosolids quality is known.

Parameters that should be analyzed are discussed briefly below (4). Although all of these may not be critical to the design of a particular disposal system, a complete analysis is necessary, because the biosolids must be adequately characterized.

(a) *Concentration*. Concentration or solids content of biosolids is related to the nature of wastewater treatment and biosolids processing steps. The type and operation of dewatering equipment may have a significant effect on the biosolids concentration. A certain degree of flexibility should be incorporated into the design of landfills to compensate for the variability in solids concentration of dewatered biosolids.

(b) *Volatile content*. Volatile solids are a measure of the organic content present in the solid fraction of biosolids. This organic matter is eventually broken down into methane gas and other digestion by-products. Typically, volatile solids represent 60% to 80% of the total solids in raw primary biosolids and 30% to 60% in anaerobically digested primary solids.

(c) *Nitrogen.* Nitrogen found in biosolids is a potential source of groundwater pollution. The total quantity and type of nitrogen are of importance. Nitrate is relatively mobile in soil and is therefore of concern.

(d) *Inorganic ions*. Inorganic ions such as heavy metals are found in most municipal biosolids. These are more readily leached if soil and biosolids are acidic. If near neutral or alkaline conditions are maintained, the metals will not be as readily leached from the biosolids or through the soil.
(e) *Bacteriological quality*. Biosolids treatment systems reduce the number of pathogens (8, 10–12) and the possibility of pathogenic contamination associated with landfilling of biosolids.
(f) *Toxic organic compounds*. Toxic organic compounds can present potential contamination problems. Solids contaminated with toxic materials must be placed in appropriately designated disposal facilities.
(g) *pH*. Acidic conditions promote leaching of heavy metals and other compounds from the biosolids.

10.2. Selection of a Landfilling Method

Relationships between the characteristics of alternative landfill sites, the characteristics of the biosolids to be landfilled, and the landfill method need to be considered in the preliminary planning process. These relationships are summarized in Table 13.6.

10.3. Site Selection

Site selection is a critical process in the planning of a biosolids landfill project. It is directly related to the method of ultimate disposal. The site finally selected must be suitable for the type of biosolids to be disposed of and situated in a convenient, yet unobtrusive, location.

Table 13.6
Biosolids and site conditions

Method	Sludge solids content, %	Appropriate sludge characteristics	Appropriate hydrogeology	Appropriate ground slope
Narrow trench	15–28	Unstabilized or stabilized	Deep groundwater and bedrock	<20 %
Wide trench	≥20	Unstabilized or stabilized	Deep groundwater and bedrock	<10 %
Area fill mound	≥20	Stabilized	Shallow groundwater or bedrock	Suitable for steep terrain as long as level area is prepared for mounding
Area fill layer	≥15	Unstabilized or stabilized	Shallow groundwater or bedrock	Suitable for medium slopes but level ground preferred
Diked containment	≥20	Stabilized	Shallow groundwater or bedrock	Suitable for steep terrain as long as a level area is prepared inside dikes
Sludge/refuse mixture	≥3	Unstabilized or stabilized	Deep or shallow groundwater or bedrock	<30 %
Sludge/soil mixture	≥20	Stabilized	Deep or shallow groundwater or bedrock	<5 %

Source: US EPA.

10.3.1. Site Considerations

The following factors must be considered during the evaluation of possible landfill sites. Information on these factors should therefore be collected and assessed in advance of the final decision making process.

(a) *Haul distance*. The most favorable haul conditions combine level terrain and minimum distances.
(b) *Site life and size*. The site life and size are directly related to the quantity and characteristics of the biosolids and the method used for landfilling. Since the entire site cannot be used as fill area, both the gross area and the usable or fill area must be considered in determining the site size. Initially, the life of the site can be estimated. As the landfill is used, the expected life should be reevaluated to ensure adequate capacity for future operations.
(c) Topography. In general, biosolids landfilling is limited to sites with minimum slopes of 1% and maximum slopes of 20%. Flat terrain tends to result in ponding, whereas steep slopes erode.
(d) *Surface water*. The location and extent of surface waters in the vicinity of the landfill site can be a significant factor in the selection process. Existing surface waters and drainage near proposed sites should be mapped and their present and proposed uses outlined. Leachate control measures including collection and treatment may be required as part of the landfill design.
(e) *Soils and geology*. Soil is an important determinant in the choice of an appropriate biosolids landfilling site. Properties such as texture, structure, permeability, pH, and cation exchange capacity, as well as the characteristics of soil formation, may influence the selection of the site. The geology of possible landfill sites should be thoroughly examined to identify any faults, major fractures and joint sets. The possibility of aquifer contamination through irregular formations must be studied.
(f) *Groundwater*. Data on groundwater in the vicinity of potential landfill sites is essential to the selection process. Knowledge of characteristics such as the depth to groundwater, the hydraulic gradient, the quality and use of the groundwater, and the location of recharge zones is essential for determining the suitability of a potential landfill site.
(g) *Vegetation.* The type and quantity of vegetation in the area of proposed landfill sites should be considered in the evaluation. Vegetation can serve as a natural buffer, reducing visual impact, odor, and other nuisances. At the same time, clearing a site of timber or other heavy vegetation can add significantly to the initial project costs.
(h) *Meteorology*. Prevailing wind direction, speed, temperature, and atmospheric stability should be evaluated to determine potential odor and dust impacts downwind of the site.
(i) *Environmentally sensitive areas*. Environmentally sensitive areas such as wetlands, flood plains, permafrost areas, critical habitats of endangered species, and recharge zones of aquifers should be avoided when selecting a landfill site (5, 6).
(j) *Archaeological and historical significance*. The archaeological and historical significance of proposed sites should be determined early in the evaluation process. Any significant finds at the selected site must be accommodated prior to final approval.
(k) *Site access*. Haul routes should be major highways, or arterials, preferably those with a minimum of traffic during normal transport hours. Proposed routes should be studied to determine impacts on local use and the potential effects of accidents. Transport through nonresidential areas is preferable to transport through residential areas, high-density urban areas, and areas with congested traffic. The access roads to the site must be adequate for the anticipated traffic loads.
(l) *Land use*. Zoning restrictions and future development on potential sites should be considered in the selection process. Ideally, the biosolids landfill site should be located on land considered unsuitable for higher uses; however, the designer should be aware that this may be a politically sensitive issue and maximum public participation must be assured.

(m) *Costs.* Cost-effectiveness of each potential landfill site must be evaluated. Factors to be included in the economic evaluation include capital costs and operating and maintenance (O&M) costs. In the latter category, biosolids hauling may prove to be a significant component. The trade-offs between high capital and high O&M costs will depend on the design life of the landfill. These trade-offs will become evident when the total annual (amortized capital and O&M costs) are compared.

10.3.2. Site Selection Methodology

The selection procedure can be roughly divided into three phases:

(a) Initial inventory and assessment of sites
(b) Screening of potential sites
(c) Final site selection

Initial inventory and assessment is designed to develop a list of potential sites that can be evaluated and rapidly screened to produce a manageable number of candidate sites. Information used in this phase is generally available and readily accessible. Investigation of each option becomes more detailed as the selection procedure progresses.

Initial assessments will consist of identifying Federal, State, and local regulatory constraints, eliminating inaccessible areas, locating potential sites, roughly assessing the economic feasibility of such sites, and performing preliminary site evaluations. The less desirable sites are eliminated on the basis of preliminary economics, regulatory, and technical information. A public participation program is initiated (33). Attitudes of the public should be determined early. The public may assist in identifying candidate sites.

Sites remaining after the initial assessment are subjected to closer scrutiny. Information used in evaluating each option is more detailed and somewhat more site-specific than in the initial assessment. Remaining sites may be rated by a scoring system including both objective and subjective evaluations. Candidate systems with lowest overall ratings are eliminated, and the higher rated systems are carried forward for final evaluation.

Site selection findings for the remaining candidate systems should provide input into an environmental impact report, if required. Public attitudes toward the remaining sites should also be determined.

Methodology for final site selection is similar to that for the screening procedure just discussed, in that rating systems are still used. However, each site remaining is investigated in greater detail. Public hearings may also be scheduled so that final inputs can be received from local government officials and the public.

Once the best sites are determined, they must be acquired. Site acquisition should begin immediately following acceptance of the program by local, State, and Federal regulatory authorities. The several acquisition procedures include: purchase option, outright purchase, lease, condemnation and/or other court action, and land dedication.

It will generally prove advantageous to purchase the site rather than hold a long-term lease. The managing agency's responsibility will normally extend well beyond the life of the site. Certain advantages may also be gained by leasing with an option to buy the site at the time of planning approval. A purchase option assures the availability of land upon completion of the facility planning process. This approach also allows time for the previous owner to gradually phase out operations, if desired.

11. FACILITY DESIGN

11.1. Regulations and Standards

Local, State, and Federal regulations and standards must be fully understood before the landfill is designed. Consideration must be given to requirements governing the degree of biosolids stabilization, the loading rates, the frequency and depth of cover, monitoring, and reporting (5, 6). The design should conform to all building codes and should include adequate buffer zones to protect public roads, private structures, and surface waters.

US EPA Rule Part 503 (5, 6) states that in case landfill sites use liners (hydraulic conductivity $\leq 1 \times 10^{-7}$ cm/s) and leachate collection system there are no pollutant concentration limits because pollutant leaching will be collected and treated. Where landfill sites are with no liners, limits on three pollutants (arsenic, chromium, and nickel) are established. While these vary based on the distance of the active landfill boundary from the site property line, the most extreme values allowed are listed in Table 13.7.

The Rule also requires that the landfill operation does not cause the maximum contaminant level for nitrates in groundwater to be exceeded or to cause the existing level of nitrates to be exceeded if it already exceeds the maximum contaminant level.

Part 503 also requires that the biosolids be either of Class A or Class B with respect to pathogen control unless the biosolids are covered daily with soil or other material. It must be stated that in many locations state regulations may be more strict even requiring a liner system.

Obtaining permits for construction and operation of biosolids landfills can be a long and costly process. To minimize delays associated with this task, permit application should be initiated early in the design stage. A sound regulatory-consultant relationship and a mutual understanding should be developed.

Table 13.7
Maximum allowable pollutant concentrations[a] in biosolids for disposal in landfills—no liner and leachate collection systems

Distance from active landfill boundary to property line (m)	Arsenic (mg/kg)	Chromium (mg/kg)	Nickel (mg/kg)
0 to <5	30	200	210
25 to <50	34	220	240
50 to <75	39	260	270
75 to <100	46	300	320
100 to <125	53	360	390
125 to <150	62	450	420
≥150	73	600	420

Source: US EPA.
[a] All pollutant concentrations are dry-weight basis.

The following is a partial list of the permits which may be required (4):

(a) US EPA special permit if landfill is in wetlands or other sensitive areas (5, 6)
(b) Army Corps of Engineers permit for construction of levees, dikes, or containment structures to be placed in the water in a wetlands area
(c) Office of Endangered Species permit if landfill is located in critical habitat of an endangered species
(d) Solid Waste Management permit
(e) Special Use permit
(f) Highway Department permit
(g) Construction permit
(h) Building permit
(i) Drainage and/or Flood Plain Alteration permit

11.2. Site Characteristics

Site characteristics should be clearly described and analyzed to ensure the suitability of the landfill site and the method of landfilling. Design phase work will build upon planning phase data but will be carried to a higher level of detail and include working drawings.

(a) **Site Plan.** The site plan should contain the following minimum information:
- Boundaries of fill area and buffer zones
- Topographic features and slopes of fill area and buffer zones
- Location of surface water, roads, and utilities
- Existing and proposed structures and access roads
- Vegetation to remain and to be removed; areas to be vegetated

(b) **Soils.** The soil characteristics at the landfill site should be thoroughly catalogued and mapped. The information of most importance to the design and operation of the landfill includes depth, texture, structure, bulk density, porosity, permeability, moisture, stability, and ease of excavation. Areas with rocky soils or extensive rock outcrops should be noted. The pH and cation exchange capacity have a direct bearing on heavy metal transport through the soil. Translocation of metals must be considered to ensure protection of surface and groundwater supplies.

(c) **Groundwater.** The groundwater aquifers underlying the landfill site must be located. Depth of the aquifer under varying conditions should be determined at several locations. Other characteristics such as the direction and rate of flow, the hydraulic gradient, the quality, and present and planned uses should also be established. Location of the primary recharge zones is critical in protecting quality.

(d) **Subsurface Geology.** The geological formations underlying the landfill are important in establishing the design parameters. Critical design parameters include the depth, distribution, and characteristics of subsurface soils in relation to stability and groundwater transmissibility.

(e) **Climate.** Climate can influence many factors in the design of landfills. Climatic conditions effect rate of organic decomposition, the composition and quantity of leachate and runoff, the day-today fill operations, and the dispersion of odors and dust. Information such as seasonal temperature, precipitation, evaporation, wind direction and speed and atmospheric stability, can be obtained from a local weather station.

(f) **Land Use.** The present and proposed use of the landfill site and adjacent properties should be evaluated. If the site is already dedicated to refuse or biosolids disposal, it is unlikely that expanding it will result in adverse impacts. However, if the site is located in or near a populated area, extensive control measures may be needed to eliminate concerns and minimize any public nuisance which would detract from the value of adjacent properties.

11.3. Landfill Type and Design

More than one biosolids landfill method may be suitable for the selected site, as shown in Table 13.2. If this is the case, a method must be selected before the final design is begun.

Maximizing utilization of the site is an important consideration in method selection. If daily cover is to be applied, the daily biosolids generation rate will affect the net capacity of the site. If several days are required to fill a trench, as the result of low biosolids generation, and cover is required each day, then the ratio of biosolids/cover will be less than for sites managing larger biosolids quantities. The net biosolids capacity will be higher at sites where trenches are filled each day.

The amount by which the net capacity of the site will be reduced will vary with the landfill methods, the specific site, and the daily biosolids generation rate. Before a final method is selected, estimates of net capacity and site life should be made for each. Additional design criteria are summarized in Table 13.8.

11.4. Ancillary Facilities

Ancillary facilities may be needed in association with the landfill site. These are described briefly in the following sections.

(a) **Leachate Controls.** Leachate from the landfill site must be contained and treated to eliminate potential water pollution and/or potential public health problems. In many cases, leachate containment and treatment may be required by state or local regulations. Numerous methods are available for controlling leachate, including drainage, natural attenuation, soil or membrane liners, or collection and treatment (34–42). The method and the design features chosen are specific for each project. Table 13.9 depicts biosolids-only leachate quality for one site sampled over 2 years.

(b) **Gas Control.** Gas produced by decomposition of organic matter is potentially dangerous. This condition is of particular concern if the landfill is located near a populated area. Methane gas, in particular, is highly explosive if confined in an enclosed area. Control of the gases produced at the landfill must be provided. Two widely accepted methods control paths of gas migration. Permeable methods usually consist of a gravel-filled trench around the fill area for intercepting migrating gas and venting it to the atmosphere. Impermeable methods consist of placing a barrier of low permeability material, such as compacted clay, around the fill area to minimize lateral movement of gas. This method provides for gas venting through the cover material. In general, methane recovery is not cost-effective at biosolids-only or small co-disposal sites.

(c) **Roads.** Paved access and on-site roads are necessary at the landfill site. Temporary roads may be constructed of well compacted natural soil or gravel. Considerations should include grades, road surface and stability, and climate. Grades in excess of 10% should be avoided. Provisions should be made to allow trucks to turn around within the site area.

(d) **Soil Stockpiles.** Storage area should be provided for on-site stockpiling of transported soils where on-site soils are insufficient or their use inappropriate. The quantity and type of soil to be stockpiled depends on the individual demands of the landfill. Stockpiles may also be desirable for winter operations where frozen ground may limit excavation.

(e) **Inclement Weather Areas.** Special landfill areas should be placed near the entrance to the site so that operations may be continued during inclement weather. Paved or all-weather roads should be provided for working these sites.

(f) **Structures.** An office and employee facilities should be located at the landfill site. For large operations, a permanent structure should be provided. At smaller sites a trailer might suffice. An equipment barn and shop may be desirable for some locations.

Table 13.8
Landfill design criteria

Method	Sludge solids content, %	Trench width, ft	Bulking required	Bulking agent	Bulking ratio[a]	Cover thickness, ft		Imported soil required	Sludge application rate, $yd^3/acre$[b]	Equipment
						Interim	Final			
Sludge only-trench fill										
Narrow trench	15–20[c]	2–3	No	-	-	-	2–3	No	1,200–5,600	Backhoe with loader, excavator, trenching machine
	20–28[c]	3–10	No	-		-	3–4			
Wide trench	20–28[c]	10	No	-	-	-	3–4	No	3,200–14,500	Track loader, dragline, scrapor, track dozer
	≥28[d]	10	No	-			4–5			
Sludge only-area fill										
Area fill mound	≥20[c,d]	-	Yes	Soil	0.5–2 soil: 1 sludge	3	3–5	Yes	3,000–14,000	Track loader, backhoe with loader, track dozer
Area fill layer	≥15[d]	-	Yes	Soil	0.25–1 soil: 1 sludge	0.5–1	2–4	Yes	2,000–9,000	Track dozer, grader, track loader
Diked containment	20–28[c]	-	No	Soil	0.25-		3–4	Yes	4,800–15,000	Dragline, track dozer, scraper
	≥28[d]		No	Soil	1 sludge					
Codisposal with refuse										
Sludge/refuse mixture	≥3[d]	-	Yes	Refuse	4–7 tons refuse: 1 wet ton sludge	0.5–1	2	No	500–4,200	Dragline, track dozer
Sludge/soil mixture	≥20[d]	-	Yes	Soil	1 soil: 1 sludge	0.5–1	2	No	1,600	Tractor with disc, grader, track loader

[a]Volume basis unless otherwise noted.
[b]In actual fill areas.
[c]Land-based equipment.
[d]Sludge-based equipment.
[e]But sometimes used.
Source: US EPA.

Table 13.9
Leachate quality from biosolids-only-landfill

Constituents	Values[b]
Constituents	
pH	6.7
TOC	1,000[c]
COD	5,100[d]
Ammonia nitrogen	198[d]
Nitrate nitrogen	0.28[d]
Chloride	6.7
Sulfate	10
Specific conductivity	3,600[e]
Cadmium	0.017
Chromium	1.1
Copper	1.3
Iron	170
Mercury	0.0004
Nickel	0.31
Lead	0.60
Zinc	5.0

[a]Data from "Site 8" monitored from July 1975 through September 1977. First received sludge in 1973. Receives unstabilized primary and WAS, gravity thickened and centrifuged. Sludge is lagooned, allowed to dry, and covered with soil. Soil characteristics: sand and gravel, glacial deposites.
[b]Specific conductivity in micromhos/cm, pH in units, all others in mg/L.
[c]Ranged from 3,000 mg/L to 1 mg/L.
[d]Limited to early part of sampling program.
[e]Ranged from 10,000 micromhos/cm 340 micromhos/cm.
Source: US EPA.

(g) **Utilities.** Electrical, water, communication, and sanitary services should be provided for large landfill operations. Chemical toilets, bottled water, and on-site electrical generation may reduce the cost of obtaining services from utility companies. This approach may be appropriate for remote sites.

(h) **Fencing.** The landfill site should be fenced. Access should be limited to one or two secured entrances. The height and type of fence should suit local conditions. A 6-ft (1.8 m) chain link fence topped with barbed wire will restrict trespassers; a wooden fence or hedge is effective for screening the operation from view, and a 4-ft (1.2 m) barbed wire fence will keep cattle or sheep away from the site area.

(i) **Lighting.** Portable lighting should be provided if landfill operations are carried out at night. Permanent lights should be installed for all structures and heavily used access roads.

(j) **Wash Racks.** A cleaning program should be required for frequently used equipment. A curbed wash pad and collection basin should be provided to contain the contaminated washwater for treatment.

(k) **Monitoring Wells.** It is crucial to monitor groundwater. The number, type, and location of monitoring wells and monitoring frequency should be designated to meet specific conditions associated with the landfill.

(l) **Landscaping.** Depending on the size and location of the landfill, landscaping may be an important design factor. The aesthetic acceptability of the landfill is critical, especially in an urban or densely populated area. In general, shrubbery chosen should require little maintenance and become an effective visual barrier.

11.5. Landfill Equipment

A wide variety of equipment may be required for a biosolids landfill. The type of equipment depends on the landfill method employed and on the quantity of biosolids to be disposed of. Equipment will be required for biosolids handling, excavation, backfilling, grading, and road construction. Table 13.10 presents typical equipment performance characteristics for various biosolids landfilling methods.

11.6. Flexibility, Performance, and Environmental Impacts

Because biosolids characteristics and quantities may change, a landfill site should be designed with maximum flexibility. Since the life of a landfill is difficult to accurately predict, expansion may be needed sooner than originally planned or it may be delayed. Any change in wastewater treatment or biosolids management processes may affect the nature and quantity of biosolids produced. Operational modifications may be needed if these changes are drastic. The landfill design should be such that changes can be made without major disruption to operations.

Reliability is another important factor in designing a landfill operation. Operation should continue even in inclement weather. Special work areas and storage facilities should be available on site for emergency operations or unexpected equipment failures.

Although the overall performance of a biosolids landfill may be difficult to predict accurately, certain operating parameters should be estimated. The site life depends on many factors; an estimate is needed for purposes of economic evaluations and future planning. Biosolids application rate and soil cover requirements should be estimated before scheduling initial operations. Performance can be more closely predicted after actual operating experience is gained.

Specific areas of environmental impact vary among landfill locations. Crucial impact areas include: traffic, land use, air quality, surface and groundwater quality, public health, aesthetics, wildlife, and habitats of endangered species. Adverse impacts should be mitigated during the site selection process or by specific measures in the design.

12. OPERATION AND MAINTENANCE

A biosolids landfill should be viewed as an ongoing construction site. Unlike conventional construction, however, the operating parameters of a biosolids landfill often change and may require innovative alterations and contingency plans. An effective landfill requires a detailed

Table 13.10
Landfill equipment performance characteristics

Landfill method	Submethod	Equipment function	Equipment type: Trenching machine	Backhoe with loader	Excavator	Track loader	Wheel loader	Track dozer[a]	Scraper	Dragline	Grader	Tractor with disc
Trench	Narrow trench	Trench construction	G	G	G	-	-	-	-	G	-	-
		Covering	G	G	F	G	G	G	-	G	-	-
	wide trench	Trench construction	-	-	-	G	F	G	G	G	-	-
		Covering	-	-	-	F	-	G	-	G	-	-
Area fill	Mound	Soil hauling	-	F	-	F	G	-	G	-	-	-
		Mixing	-	F	-	G	F	G	-	-	-	-
		Sludge hauling	-	F	-	F	G	-	F	-	-	-
		Mounding	-	G	-	G	F	F	-	-	-	-
		Covering	-	F	-	G	F	G	-	G	-	-
	Layer	Soil hauling	-	F	-	F	G	-	G	-	-	-
		Mixing	-	F	-	G	G	G	-	-	-	-
		Sludge hauling	-	F	-	F	G	-	F	-	-	-
		Layering	-	-	-	F	-	G	G	-	G	-
		Covering	-	-	-	F	-	G	G	-	G	-
	Diked containment	Soil hauling	-	F	-	F	G	-	G	-	-	-
		Dike construction	-	-	-	F	F	G	G	-	-	-
		Covering	-	-	-	-	-	G	-	G	-	-
Codisposal	Sludge/refuse	Spreading	-	-	-	F	-	G	-	-	-	-
		Covering	-	-	-	F	-	G	F	-	-	-
	Sludge/soil	Sludge spreading	-	-	-	F	-	G	-	-	F	-
		Mixing	-	-	-	-	-	F	-	-	-	G
		Hauling	-	-	-	G	F	-	F	-	-	-
		Covering	-	-	-	F	F	G	F	-	-	-

Legend
G = Good. Fully capable of performing function listed. Equipment could be selected solely on basis of function listed.
F = Fair. Marginally capable of performing function listed. Equipment should be selected on basis of full capabilities in other function.
- = Not applicable. Cannot be used for function listed.

[a] Caterpillar D-6 generally is the largest track doser appropriate for a sludge landfill although some engineers are investigating the use of the Caterpillar LG-T, double-wide track doser.

Source: US EPA.

operational plan. Equipment selection should be compatible with biosolids characteristics, site conditions, and landfill method.

Operational procedures can be separated into those specific to the landfill method and those applicable to biosolids landfills in general. Method-specific procedures include: site preparation, biosolids unloading, biosolids management and covering. General procedures include scheduling, equipment selection and maintenance, management and reporting, safety, and environmental controls. These procedures are discussed in detail in Municipal Biosolids Landfill (30) and in Sanitary Landfill Design and Operation (31). Important points are summarized below.

12.1. Operations Plan

As with any construction activity, biosolids landfilling must proceed according to detailed plans and operating schedules. The operation plan should address all relevant method-specific or general operating procedures for the landfill, including:

(a) Hours of operation
(b) Measuring procedures
(c) Traffic flow and unloading procedures
(d) Special wastes handling
(e) Inclement weather operations
(f) Environmental monitoring and control practices

An operations plan is an important tool for providing continuity of activities, monitoring and control of progress, and personnel training.

12.2. Operating Schedule

Major features of the operating schedule include:

(a) Hours of operation
(b) Availability of qualified personnel
(c) Site preparation schedules
(d) Equipment maintenance schedules

The hours of operation must be such that the site is open when biosolids is to be received. If variations in the rate of receipt are expected during the day, it may be desirable to schedule for equipment and personnel accordingly. The schedule may need to provide for the application of daily soil cover.

12.3. Equipment Selection and Maintenance

Equipment selection depends largely upon the landfill method, design dimensions, and biosolids quantity. Selection must be based upon the functions to be performed and the cost of alternate machines. Table 13.10 summarized general selection criteria. Table 13.11 presents examples of equipment choices for seven landfill schemes.

Equipment maintenance can be more expensive than the amortized annual purchase cost. A scheduled preventive maintenance program should be followed to control maintenance costs.

Table 13.11
Typical equipment type and number as a function of landfill method and site loading

	Trench method											Area fill method															Codisposal method[a]									
	Narrow trench					Wide trench						Mound					Layer					Diked containment						Sludge/refuse					Sludge/soil				
Equipment	1[b]	2[c]	3[d]	4[e]	5[f]	1	2	3	4	5		1	2	3	4	5	1	2	3	4	5	1	2	3	4	5		1	2	3	4	5		1	2	3	4	5
Trenching machine				1	2																															
Backhoe with loader	1	1		1[g]	1							1[g]	1[g]	1[g]	1																					
Excavator			1																																	
Track loader						1	1[g]	1		1[g]		1	1	1	1	1														1			1[g]	1[g]	2[g]
Wheel loader															1	1					1[g]															
Track dozer		1[g]	1	1	2[g]		1[g]	1	1	2[g]				1[g]	1	1	1	1	1	2[g]	2	1	1[g]	1[g]	1	2[g]			1[g]	1	1					
Scraper									1[g]	1				1[g]	1[g]	1		1[g]	1[g]	1[g]	1			1[g]	1	1										
Draglino																							1	1	1	1										
Grader																																		1[g]	1
Tractor with disc																																1	1	1[g]	2[g]	2
Total	1	2	2	3	5	1	2	2	2	4		1	2	4	5	5	1	2	2	3	4	1	2	3	3	4		-	-	1	1	2		1	1	2	4	5

[a] Additional equipment only.
[b] Scheme 1 - 10 wet t/day.
[c] Scheme 2 - 50 wet t/day.
[d] Scheme 3 - 100 wet t/day.
[e] Scheme 4 - 250 wet t/day.
[f] Scheme 5 - 500 wet t/day.
[g] May not receive 100 percent utilization.
Source: US EPA.

Operators should perform routine daily maintenance (e.g., check fluid levels, cleaning, etc.). The operating schedule should provide periods for thorough maintenance.

12.4. Management and Reporting

Management and reporting activities include the maintenance of activity records, performance records, required regulatory reports, cost records, on-site supervision and public relations activities (43–48). Activity records include equipment and personnel accounts, biosolids, and (if applicable) solid waste receipts, cover material quantities and used site area layouts. These records become bases for scheduling site development, gauging efficiency, and any billing as required.

Performance records may be required as a part of the regulatory process. Regulatory agencies may perform periodic inspections on a scheduled or an unscheduled basis. Operating and supervisory personnel must be aware of these requirements.

For the purposes of safety and control, the site should be staffed with two or more persons. At smaller sites, where only one operator is required, daily visits or phone checks should be made.

12.5. Safety

Providing a safe working environment at the landfill site should be a part of general O&M, and certain safety features should be built into the design. Certain practices must be followed daily to provide safe working conditions. The operations plan should have a separate safety section, as well as specific safety guidelines for each operation and feature of the landfill.

(a) **Soil and Fill Stability.** The stability of the soil and fill can present a critical safety problem, particularly with the use of large equipment. Disturbed and filled areas should be approached cautiously as should muddy areas or areas subject to erosion.

(b) **Equipment Operation.** The operation of large, earth-moving equipment presents the potential for accidents. Only fully trained operators should be allowed to use such equipment. Regular maintenance and safety checks can greatly reduce the number of accidents associated with equipment failure and operator error.

(c) **Gas Control.** Caution must be used when dealing with gas control equipment. The O&M manual should contain a complete set of instructions on the safe servicing of gas control and monitoring equipment, and the operation of this equipment should be explained periodically at operation and safety training sessions.

12.6. Environmental Controls

The protection of the environment and public health are important aspects of the landfill operation. The operations plan should contain guidelines for providing this protection and actual operations should conform to the guidelines. Potential environmental problems and the requirements for their control are summarized in Table 13.12. Critical areas are discussed below.

(a) **Environment.** Environmental protection is generally focused on leachate and runoff controls for preventing surface and groundwater contamination. Trench liners must be kept intact during

Table 13.12
Potential environmental problems and control practices

Control practice	Environmental problems									
	Spillage	Siltation and erosion	Mud	Dust	Vectors	Odors	Noise	Aesthetics	Health	Safety
Safety program										x
Maintain washrooms for personnel									x	
Training of new personnel	x	x	x	x	x	x	x	x	x	x
Use safety clamps on truck tailgates	x									x
Maintain road markings and trench barriers	x									x
Maintain fencing								x		x
Apply insecticide					x				x	
Maintain buffer areas and grass		x	x	x		x	x	x		
Proper equipment maintenance	x						x			x
Spray water/oil/liquid asphalt			x	x						
Truck wash pad (to clean trucks)			x	x		x				
Maintain grass waterways, diversion ditches, rip rap		x	x					x		
Final grading of disturbed areas		x						x		
Revegetation of disturbed areas		x	x	x				x		
Chemical masking agent						x				
Lime on site	x				x	x			x	x
Workers supplied with aerators				x		x			x	x
Cover sludge daily					x	x		x	x	x
Water diverted away from site		x	x							

Source: US EPA.

and after tilling operations (49). Drainage systems should be checked to see that they are functioning as designed. If monitoring indicates that adverse environmental impacts are occurring or pending, immediate corrective action should be taken.

(b) **Public Health.** Protection of public health should be a foremost concern in the operation of biosolids landfills. Protection of water supplies and particularly water aquifers is an obvious responsibility. In addition, control of potential disease by reduction of vectors, the adequate venting of explosive or toxic gases and the restriction of access to the landfill site are the responsibility of the operators.

(c) **Social Welfare.** Minimizing the negative aesthetic impacts of a biosolids landfill can greatly increase public acceptance. Control of odors, noise, and other nuisances is generally straightforward and should be accomplished as part of the daily operating routine. Efforts should be made to reduce the undesirable social impacts of the fill operation.

13. SITE CLOSURE

In closing a biosolids landfill site, certain criteria must be met to make the site publicly acceptable. These criteria are established according to the type of landfill and the location,

size, and ultimate use of the site. The procedures for site closure should be included in the operations manual and updated or modified as the original landfill plan is altered.

13.1. Ultimate Use

The ultimate use of the site should be described and illustrated in the O&M manual or in a separate document describing the closure of the site. The actual work involved in completing the site will depend on its ultimate use and on the care taken in day-to-day fill operations.

13.2. Grading at Completion of Filling

When each section of the landfill is completed, the final cover should be graded according to a predetermined plan. It is imperative that no biosolids become or remain exposed after the grading has been completed (50, 51).

Final grading of the site is to be performed after sufficient time has elapsed to allow for initial settlement. The final grading plan should be designed in accordance with the intended ultimate use of the landfill site. It is important that all biosolids be completely covered to the specified depth with cover material.

13.3. Landscaping

The landscaping plan should reflect the intended ultimate use of the landfill site. Where practical, landscaping may be done on completed sections before the entire fill project is completed (52).

13.4. Continued Leachate and Gas Control

Since decomposition of the organics in the biosolids may continue even after the landfill has been completed, an ongoing monitoring and control program must be maintained. Leachate and gas must be controlled even after the filling operations have stopped. The completion plans should clearly outline this program.

14. COSTS OF BIOSOLIDS DISPOSAL ON LAND (LANDFILL)

14.1. General

(a) Most biosolids disposal systems have at least four definable components: storage, collection, haul, and disposal.

(b) Treatment of biosolids is related to reducing the volume to a minimum before transporting. Typical unit processes used for volume reduction may include digestion, centrifugation, vacuum filtration, and drying beds (12–15, 53). Costs associated with these processes are not considered to be part of biosolids hauling or landfilling but are very important in the overall biosolids handling train.

(c) Storage costs are site-specific and depend largely upon the method selected in the biosolids handling train (44, 45). They may be simply the costs associated with the purchase of bins for storage of secondary or primary biosolids, a dump truck for storage of digested biosolids that have been centrifuged or vacuum filtered, or the cost associated with drying beds.

(d) Collection costs are dependent upon a time-labor relationship to transfer the biosolids from storage to the transporting vehicle, as a dump or tank truck (54). There may not be a collection cost associated with labor; however, a cost would be incurred to provide a vehicle during the

loading period. Larger facilities may require that a driver be assigned to the vehicle during loading periods. Collection costs may be significant when it is necessary to shovel biosolids from drying beds into trucks for transportation to the landfill. Collection costs are site and system specific.

(e) Transportation costs are associated with such parameters as truck cost, truck size, haul time, labor, and operating costs per unit time for items such as depreciation, fuel, insurance, maintenance, etc. (44). Operating costs may be estimated from manufacturer's rating information and used in conjunction with estimates of biosolids production from various wastewater treatment processes.

(f) Disposal costs are related to the operation and management of the final disposal facility. This cost should be minimal if the facility will integrate ultimate biosolids disposal with the disposal of refuse. When this is possible, the disposal costs may only include the costs of unloading and a landfill fee. On the other hand, if the landfill is to receive only waste biosolids; costs may be very significant as other equipment for operation of the landfill will be required. The equipment used for landfill operation may include units for excavation, placing, covering, and compaction of fill.

(g) The lowest possible moisture content attainable at a reasonable cost should be produced for economical biosolids hauling and landfill operations. A reduction of moisture content will produce a savings in storage, initial equipment, operating, and labor costs.

14.2. Hauling of Biosolids

14.2.1. Required Input Data

(a) Average wastewater flow, MGD
(b) Biosolids volume, gal/MG
(c) Raw biosolids concentration, %
(d) Dewatered biosolids concentration, %
(e) Vehicle loading time, h
(f) Round-trip haul time, h
(g) Vehicle capacity, yd^3
(h) Solids capture in dewatering process, %
(i) Distance to disposal site, mile

14.2.2. Design Parameters

(a) Biosolids volume/MG treated (Table 13.13)
(b) Biosolids concentration, 1.5% to 15% (Table 13.13)
(c) Cake concentration, 6% to 60% (Table 13.14)
(d) Vehicle capacity, yd^3/truck
(e) Truck loading time, 0.5 to 2.0 h
(f) Haul time, local conditions, h
(g) Daily work schedule, 6 to 8 h
(h) Solids capture, 70% to 99% (Table 13.14)

14.2.3. Design Procedure

(a) Compute the biosolids volume hauled, yd^3/day

$$V_B = (Q)\ (BF)\ (SS)/(CSS)\ (7.48)\ (27) \quad (1)$$

where

V_B = volume of biosolids, yd^3/day
Q = wastewater flow, MGD
BF = biosolids flow, gal/MG (Table 13.13)
SS = suspended solids in biosolids flow, % (Table 13.13)
CSS = cake suspended solids, % (14)

(b) Calculate the number of vehicles for collection and hauling of the biosolids

$$N = (V_B)\ (LT + HT)/(HPD)\ (CAP) \tag{2}$$

where

N = number of trucks required
LT = loading time, h (0.5 to 2.0 h)
HT = round-trip haul time, h (local conditions)
HPD = work schedule, h/day (6 to 8)
CAP = vehicle capacity, yd^3/truck (3 to 12)

(c) Compute the tons of biosolids hauled per day

$$TBH = (Q)\ (BF)\ (SS)\ (SCAP)(8.34)/(100)\ (CSS)\ (2000) \tag{3}$$

Table 13.13
Normal quantities of biosolids produced by various treatment processes

Wastewater treatment process	gal sludge/ MG treated	Solids %	Sludge specific gravity
Primary sedimentation			
Undigested	2950	5.0	1.02
Digested in separate tanks	1450	6.0	1.03
Trickling filter	745	7.5	1.025
Chemical precipitation	5120	7.5	1.03
Primary sedimentation and activated sludge			
Undigested	6900	4.0	1.02
Digested in separate tanks	2700	6.0	1.03
Activated sludge			
Waste sludge	19,400	1.5	1.005
Septic tanks, digested	900	10.0	1.04
Imhoff tanks, digested	500	15.0	1.04

Source: US EPA.

Table 13.14
Process efficiencies for dewatering of wastewater biosolids

Unit process	Solids capture (%)	Cake Solids (%)
Centrifugation		
Solid bowl	80–90	5–13
Disc-nozzle	80–97	5–7
Basket	70–90	9–10
Dissolved air flotation	95	4–6
Drying beds	85–99	8–25
Filter press	99	40–60
Gravity thickener	90–95	5–12
Vacuum filter	90+	28–35

Source: US EPA.

where

TBH = tons of biosolids hauled per day, t/day
SCAP = solids capture, % (Table 13.14)

14.2.4. Output Data

(d) Volume of biosolids to be dewatered, gpd
(e) Initial moisture content, %
(f) Final moisture content, %
(g) Volume of biosolids hauled, yd^3/day
(h) Truck capacity, yd^3
(i) Time to make one load, h
(j) Work schedule, h/day
(k) Number of trucks required
(l) Tons of biosolids hauled per day, t/day
(m) Distance to disposal site, mile

14.3. Energy Requirements

Actual fuel consumption varies considerably with specific biosolids, site, and operating conditions (32). Fuel consumption rates for some typical construction equipment performing light to medium work are given in Table 13.15. One case study that used biosolids, landfill operation was estimated to consume 700,000 BTU/day/t of biosolids (1 gal diesel fuel is equivalent to 140,000 BTU).

Table 13.15
Fuel energy consumption rates for some typical construction equipment

Equipment	Average diesel fuel (gal/h)	Equipment	Average diesel fuel (gal/h)
Caterpillar D-6	5.2	Grader - 25,000 lb	4.4
Caterpillar D-8	10.8	28,000 lb	4.8
Excavator - $^1/_2$ yd^3	3.4	30,000 lb	5.2
1 yd^3	5.0	40,000 lb	6.0
$1^1/_4$ to $1^1/_2$ yd^3	8.8	54,000 lb	7.9
$1^1/_2$ to 2 yd^3	11.1		
Wheel loader $1^1/_2$ yd^3	3.0	Track loader - 1 yd^3	2.4
2 yd^3	3.7	$1^1/_2$ yd^3	3.4
3 yd^3	4.6	2 yd^3	4.2
4 yd^3	6.2	2.5 yd^3	5.7
5 yd^3	9.0	3 yd^3	7.4
7 yd^3	13.2	4 yd^3	11.3
Tractor-scraper, small	4.9		
Medium	11.4		
Targe	15.8		

Source: US EPA.

14.4. Costs

Construction and O&M costs for the two trench and three area fill methods are shown in Figures 13.1 to 13.4. All costs are in 1978 dollars (Cost Index = 235.78). To obtain the values in terms of the present 2008 USD using the Cost Index for Utilities shown in Appendix, multiply the costs by a factor of 2.34 (29). Also, take notice of the following items (32):

(a) Site and equipment costs include land [USD 2,500/acre (2008 USD 5,850/acre)], site preparation (clearing, grubbing, surface water control ditches and ponds, monitoring wells, soil stockpiles, roads, and facilities), equipment purchase, engineering (6%). Actual fill area consumes 50% of total site area.
(b) Operating costs include labor (USD 8/h [2008 USD 19/h], including fringe, overhead, administration), equipment fuel, maintenance, and parts; utilities; laboratory analysis of water samples; supplies and materials.
(c) Actual costs vary considerably with specific biosolids and site conditions (28, 32, 55, 56).

15. EXAMPLES

15.1. Example 1. Typical Biosolids Application Rate Scenario

The recommended minimum amount of nitrogen needed by a typical corn crop to be grown in New Jersey is 120 lb/acre/year. Biosolids containing 3% nitrogen could be applied at up to 5.4 dry ton/acre if used to supply all the nitrogen needed by the crop (i.e., no other

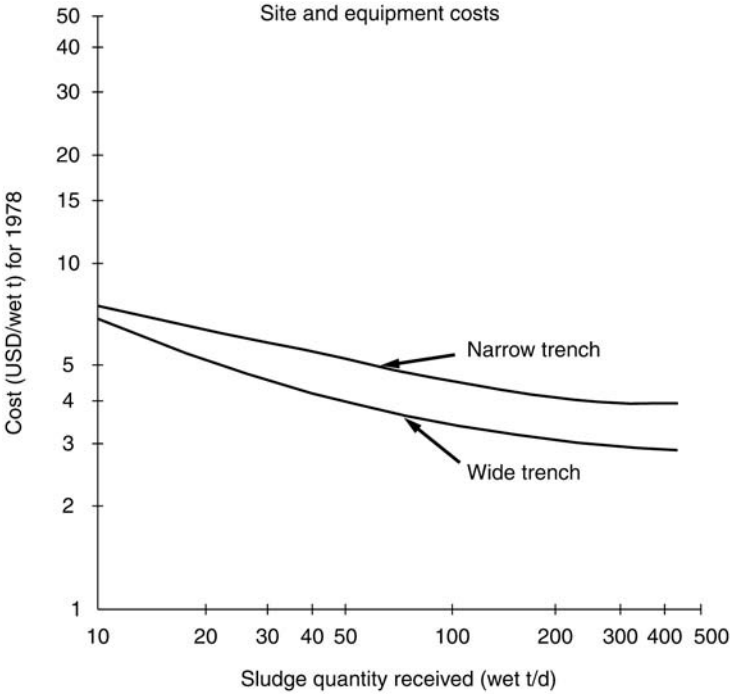


Fig. 13.1. Site and Equipment costs for narrow and wide trench landfill. *Source*: US EPA.

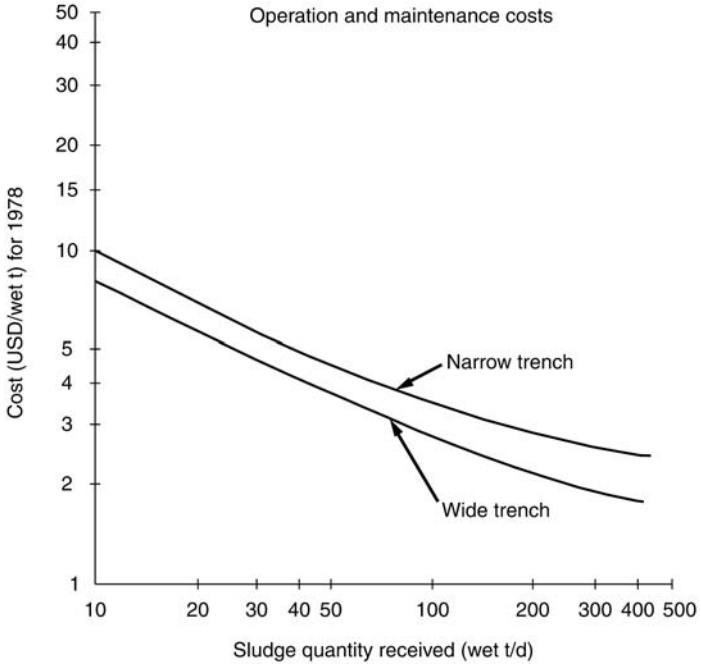


Fig. 13.2. Operation and maintenance costs for narrow and wide trench landfill. *Source*: US EPA.

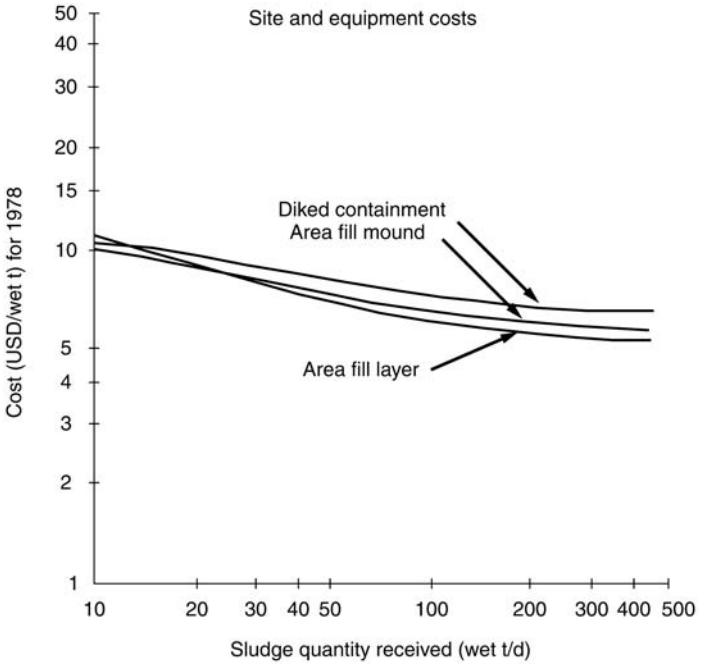


Fig. 13.3. Site and equipment costs for the three types of area landfill. *Source*: US EPA.

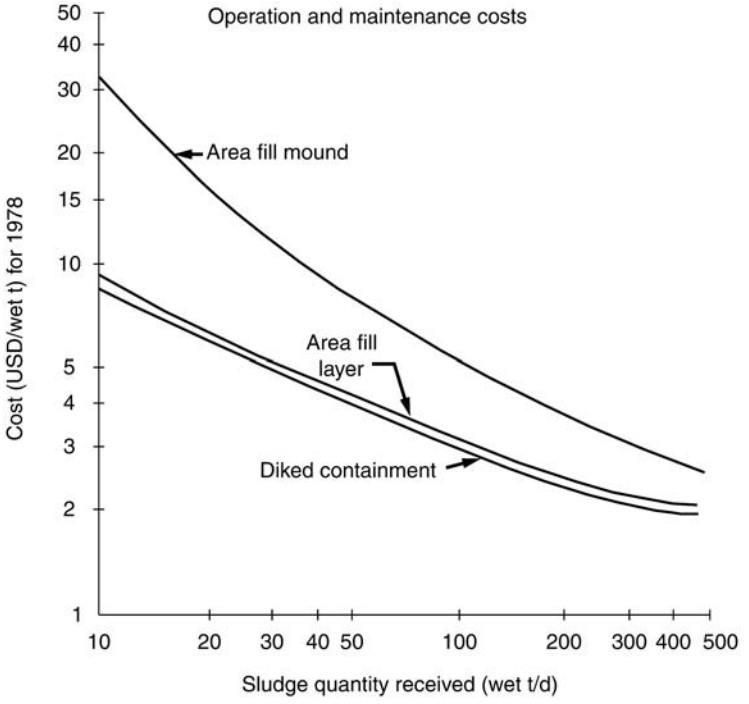


Fig. 13.4. Operation and maintenance costs for the three types of area landfill. *Source*: US EPA.

nitrogen fertilizers used). A city producing 10 t/day of dry biosolids would require access to almost 700 acres of corn. If the biosolids contained only 1.5% nitrogen, twice as many tons could be applied per acre, requiring only half as many acres to land apply the same amount of biosolids generated.

15.2. Example 2. Hauling of Biosolids

A 1.0 MGD wastewater treatment plant is planned to dispose of its biosolids in a landfill. The plan had the following design parameters:

(a) Biosolids production volume = 2700 gal/MG of wastewater flow
(b) Biosolids suspended solids content = 6%
(c) Biosolids cake solids content = 12%
(d) Solids capture = 95%
(e) Hauling truck capacity = 4 yd^3
(f) Truck loading time = 1.5 h
(g) Transport driving time to landfill = 2.5 h

Compute the following:

(a) Volume of biosolids to be hauled
(b) Number of vehicles required
(c) Tons of biosolids hauled per day

Solution

(a) Volume of biosolids to be hauled

$$V_B = (Q)\ (BF)\ (SS)/(CSS)\ (7.48)\ (27) \tag{1}$$

where
V_B = biosolids volume hauled, yd^3/day
Q = wastewater flow, 1.0 MGD
BF = biosolids flow, 2700 gal/MG
SS = suspended solids in biosolids flow, 6%
CSS = cake suspended solids, 12%

$$V_B = (1.0)(2700)(6)/(12)(7.48)(27)$$

$$V_B = 6.7\ yd^3/day$$

(b) Number of vehicles required

$$N = (V_B)\ (LT + HT)/(HPD)\ (CAP) \tag{2}$$

where
N = number of trucks required
V_B = biosolids volume hauled, 6.7 yd^3/day
LT = loading time, 1.5 h
HT = round-trip haul time, 2.5 h
HPD = work schedule, 8 h/day

CAP = vehicle capacity, 4 yd^3/truck

$$N = 6.7(1.5 + 2.5)/(8)(4)$$

$$N = 0.84$$

Therefore use 1 truck at 2 trips/day

(c) Tons of biosolids hauled per day

$$\text{TBH} = (Q)\ (\text{BF})\ (\text{SS})\ (\text{SCAP})\ (8.34)/(100)\ (\text{CSS})\ (2000) \tag{3}$$

where
TBH = tons of biosolids hauled per day, t/day
Q = wastewater flow, 1.0 MGD
BF = biosolids flow, 2700 gal/MG
SS = suspended solids in biosolids flow, 6%
SCAP = solids capture, 95%
CSS = cake suspended solids, 12%

$$\text{TBH} = (1.0)(2700)(6)(95)(8.34)/(100)(12)(2000)$$

$$\text{TBH} = 5.3\ \text{t/day}$$

NOMENCLATURE

BF = biosolids flow, gal/MG
CAP = vehicle capacity, yd^3/truck
CSS = cake suspended solids, %
HPD = work schedule, h/day
HT = round-trip haul time, h
LT = loading time, h
N = number of trucks required
Q = wastewater flow, MGD
SCAP = solids capture, %
SS = suspended solids in biosolids flow, %
TBH = tons of biosolids hauled per day, t/day
V_B = volume of biosolids, yd^3/day

REFERENCES

1. US EPA, *Land Application of Biosolids*, Biosolids Technology Fact Sheet, EPA 832-F-00-064, US Environmental Protection Agency, Office of Water, Washington, DC (2000).
2. NSFCH, *Biosolids Recycling: Benefit Technology for a Better Environment*, WWBLGN59, National Small Flows Clearinghouse, West Virginia University, Morgantown, WV (1994).
3. US EPA, *National Pretreatment Program: Report to Congress*, EPA 21 W-4004, US Environmental Protection Agency, Washington, DC (1991).

4. US EPA, *Process Design Manual for Sludge Treatment and Disposal*, EPA625/1-79-001, US Environmental Protection Agency, Washington, DC (1979).
5. US EPA, *Standards for the Use or Disposal of Sewage Sludge* (*40 Code of Federal Regulations Part 503*), US Environmental Protection Agency, Washington, DC (1993).
6. US EPA, *Amendments to the Standards for the Use or Disposal of Sewage Sludge* (*40 Code of Federal Regulations Part 503*), US Environmental Protection Agency, Washington, DC (1995).
7. US EPA, *A Guide to the Biosolids Risk Assessments for the EPA Part 503 Rule*, EPA832-B-93-005, US Environmental Protection Agency, Washington, DC (1995).
8. US EPA, *Control of Pathogens and Vector Attraction in Sewage Sludge*, Environmental Regulations and Technology, US Environmental Protection Agency, Washington, DC (1992).
9. US EPA, *Environmental Regulations and Technology, Use and Disposal of Municipal Wastewater Sludge*, EPA 625/10-84-003, US Environmental Protection Agency, Cincinnati, OH (1984).
10. K. Arnold, J. H. Dunn, and D. Sievers, *Biosolids Standards for Pathogens and Vectors*, Water Quality Initiative publication, WQ424, Web http://muextension.missouri.edu/explore/envqual/wq0424.htm (2004).
11. N-Viro, *How Biosolids Are Regulated*, N-Viro International Corporation, http://www.nviro.com/regs.htm (2004).
12. L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.), *Biosolids Treatment Processes*, Humana Press, Totowa, NJ (2007).
13. US EPA, *Centrifugal Dewatering/Thickening*, Biosolids Technology Fact Sheet, EPA 832-F-053, US Environmental Protection Agency, Office of Water, Washington, DC (2000).
14. US EPA, *Belt Filter Press*, Biosolids Technology Fact Sheet, EPA 832-F-00-057, US Environmental Protection Agency, Office of Water, Washington, DC (2000).
15. US EPA, *Filter Press, Recessed Plate,* Biosolids Technology Fact Sheet EPA 832-F-00-058, US Environmental Protection Agency, Office of Water, Washington, DC (2000).
16. US EPA, *Alkaline Stabilization of Biosolids*, Biosolids Technology Fact Sheet, EPA 832-F-00-052, US Environmental Protection Agency, Office of Water, Washington, DC (2000).
17. US EPA, *Biosolids Recycling: Beneficial Technologies for a Better Environment,* EPA 832-R-94-009, US Environmental Protection Agency, Office of Water, Washington, DC (1994).
18. US EPA, *Interagency Policy on Beneficial Use of Municipal Sewage Sludge*, US Environmental Protection Agency, Washington, DC (1981).
19. WEF, *National Outlook—State Beneficial Use of Biosolids Activities*, Water Environment Federation, Washington, DC (1997).
20. WEF Web http://www.wef.org/doc/bioquotes.html, Water Environment Federation, September (1998).
21. US EPA, Odor Management in Biosolids Management, Biosolids Technology Fact Sheet, EPA 832-F-00-067, US Environmental Protection Agency, Office of Water, Washington, DC (2000).
22. W. E. Sopper, E.M. Seaker, and R. K. Bastian (Eds), *Land Reclamation and Biomass Production and Municipal Wastewater and Sludge*, The Pennsylvania State University Press, University Park, PA (1982).
23. US EPA, *Sewage Sludge Management Primer, Technology Transfer Series*, US Environmental Protection Agency, Cincinnati, OH (1986).
24. US EPA, *Process Design Manual Land Application of Municipal Sludge* (EPA 625/1-83-016), US Environmental Protection Agency, Cincinnati, OH (1983).
25. C. Lue-Hing., D. R Zenz., and R. Kuchenrither (eds.), *Municipal Sewage Sludge Management: Processing, Utilization and Disposal*, Technomic Publishing Company, Lancaster, PA (1992).

26. L. Spinosa and P. A. Vesilind (eds.), *Sludge into Biosolids: Processing, Disposal, Utilization*, IWA Publishing, London, December (2001).
27. Metcalf and Eddy, *Wastewater Engineering Treatment and Reuse*, 4th ed, McGraw-Hill, New York (2003).
28. R. G. O'Dette, Determining the most cost effective option for biosolids and residuals management, *Proceedings of the 10th Annual Residuals and Biosolids Management Conference: 10 Years of Progress and a Look Toward the Future*, Water Environment Federation, Alexandria (1996).
29. US ACE, *Civil Works Construction Cost Index System Manual*, 110-2-1304, US Army Corps of Engineers, Washington, DC, PP 44, http://www.nww.usace.army.mil/cost, (2008).
30. US EPA, *Process Design Manual: Municipal Sludge Landfills*, EPA-625/l-78-010, SW-705, Environmental Research Information Center, Office of Solid Wastes, Cincinnati, OH, October (1978).
31. US EPA, *Disposal of Sewage Sludge into a Sanitary Landfill*, SW-71d, Office of Solid Wastes, Washington, DC (1974).
32. US EPA, *Innovative and Alternative Technology Assessment Manual*, EPA430/9-78-009, US Environmental Protection Agency, Washington, DC (1980).
33. US EPA, *Regulations on Public Participation in Programs Under the Resource Conservation and Recovery Act,* 40 CFR 25, 44 CFR 10292, The Safe Drinking Water Act, and The Clean Water Act, Office of Waste and Hazardous Materials, Washington, DC, February (1979).
34. B. Calli et al., Investigation of variations in microbial diversity in anaerobic reactors treating landfill leachate, *Water Science and Technology*, 48, 4, 105–112 (2003).
35. C. Yangin et al., A new process for the combined treatment of municipal wastewaters and landfill leachates in coastal areas, *Water Science and Technology*, 46, 8, 111–118 (2002).
36. D. Geenens et al., Combined ozone-activated sludge treatment of landfill leachate, *Water Science and Technology*, 44, 2–3, 359–365 (2001).
37. P. M. Petrangeli et al., Kaolinite sorption of Cd, Ni and Cu from landfill leachates: influence of leachate composition, *Water Science and Technology*, 44, 2–3, 343–350 (2001).
38. I. L. Yamakawa and J. Terry, Effects of aging on leachate characteristics of alkaline stabilized biosolids, *14th Annual Residuals and Biosolids Management Conference*, February/March (2000).
39. J. Doyle et al., Exceptionally high-rate nitrification in sequencing batch reactors treating high ammonia landfill leachate, *Water Science and Technology*, 43, 3, 315–322 (2001).
40. J. G. Henry and D. Prasad, Anaerobic treatment of landfill leachate by sulfate reduction, *Water Science and Technology* 41, 3, 239–246 (2000).
41. Z. R. M Ramirez et al., Treatment of landfill leachates by comparing advanced oxidation and coagulation-flocculation processes coupled with activated carbon adsorption, *Water Science and Techn*ology, 41, 1, 231–235 (2000).
42. D. D. Kozu and S. K. Lieh, Assessing denitrification rate limiting factors in a constructed wetland receiving landfill leachate, *Water Science and Technology*, 40, 3, 75–81 (1999).
43. K. M. Payne, M. Owens, and R. Wheadon, Handling and disposal alternatives for WTP residuals for the metropolitan water district of Salt Lake and Sandy, *Biosolids 2000: Building Public Support*, February (2001).
44. G. Leboucher, P. Fernandes, G. Coriton, and E. Guibelin, Options for biosolids utilization and sludge disposal: Storage and transportation, *Water Intelligence Online* (2002).
45. W. R. Uhte, Wastewater solids storage basins: a useful buffer between solids stabilization and fina1disposal, Presented at the *48th Annual Conference of the California Water Pollution Control Association*, Lake Tahoe, California, April (1976).

46. US EPA, *Subsurface Disposal of Municipal Wastewater Treatment Sludge*, SW-167c, Office of Solid Wastes, Washington, DC (1978).
47. C. J. Banks and S. Heaven, Options for biosolids utilization and sludge disposal: landfilling, *Water Intelligence Online* (2002).
48. P. Matthews, A millennium perspective on biosolids and sludge management, *14th Annual Residuals and Biosolids Management Conference*, February/March (2000).
49. M. C. L. Irene et al., Risk assessment using stochastic modeling of pollutant transport in landfill clay liners, *Water Science and Technology*, 39, 10–11, 337–341 (1999).
50. D. Leffler, C. Drill, and J. D. Oneil, A novel alkaline biosolids product as alternative landfill cover, *14th Annual Residuals and Biosolids Management Conference*, February/March (2000).
51. D. Leffler, C. Drill, and J. D. Oneil, A novel alkaline biosolids product as alternative landfill cover, *Odors and VOC Emissions 2000*, April (2000).
52. K. Fanfoni and L. M. Naylor, Beneficial reuse of biosolids in landfill closure, *Biosolids 2000: Building Public Support*, February (2001).
53. P. Brautlecht and S. Gredigk, Concept for an interlinked system of a sludge drying facility and a landfill for residual waste, *Water Science and Technology*, 38, 2, 119–125 (1998).
54. R. M. Clark and B. P. Helms, Fleet selection for solid waste collection systems, *Journal, Environmental Engineering Division, American Society of Civil Engineers*, l, 98, 71 (1972).
55. R. P. Dominak and L. A. Stone, Residuals disposal costs—a detailed analysis, *WEFTEC 2002 Conference Proceedings*, September (2002).
56. L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.), *Biosolids Engineering and Management*, Humana Press, Totowa, NJ (2008).

APPENDIX

US Yearly Average Cost Index for Utilities US Army Corps of Engineers (29)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

14
Deep-Well Injection for Waste Management

Nazih K. Shammas, Charles W. Sever, and Lawrence K. Wang

CONTENTS

Abstract Man-made or produced fluids (liquids, gases or slurries) can move into the pores of rocks by the use of pumps or by gravity. Injection well technology can predict the capacity of rocks to contain fluids and the technical details to do so safely. Underground wastewater disposal and storage by well injection is being used by both industries and municipalities to help solve environmental problems. When wells are properly sited, constructed, and operated, underground injection is an effective and environmentally safe method to dispose of wastes. Issues discussed in this chapter include regulations for managing injection wells, basic well designs, evaluation of a proposed injection well site, ways to prevent, detect, and correct potential hazards, economic evaluation of a proposed injection well system, use of injection wells for wastewater management, use of injection wells for hazardous wastes management, protection of usable aquifers, case studies of deep well injection, and practical examples.

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_14 © Humana Press, New York, NY 2009

Key Words Hazardous waste disposal • injection well • wastewater disposal • well design • case studies and examples.

1. INTRODUCTION

Underground injection is the technology of placing fluids underground, in porous formations of rocks, through wells or other similar conveyance systems (1, 2). Although rocks such as sandstone, shale, limestone appear to be solid, they can contain significant voids or pores that allow water and other fluids to fill and move through them. Man-made or produced fluids (liquids, gases or slurries) can move into the pores of rocks by the use of pumps or by gravity. The fluids may be water, wastewater, or water mixed with chemicals. Injection well technology can predict the capacity of rocks to contain fluids and the technical details to do so safely.

Underground wastewater disposal and storage by well injection is being used by both industries (3–9) and municipalities (3, 4, 6, 10–13) to help solve environmental problems. Facilities across the United States (US) discharge a variety of hazardous and nonhazardous fluids into more than 400,000 injection wells (2). A national goal established by Congress in the Clean Water Act of 1972 was to cease all discharges of pollutants to navigable waters by 1985 (14). Although treatment technologies exist, it would be very costly to treat and release to surface waters the billions and trillions of gallons of wastes that industries produce each year. Agribusiness and the chemical and petroleum industries all make use of underground injection for waste disposal. When wells are properly sited, constructed, and operated, underground injection is an effective and environmentally safe method to dispose of wastes.

Subsurface disposal of liquid wastes into aquifers is based on the concept that such wastes can be injected through wells into confined geologic strata not having other uses, thereby providing long-term isolation of the waste material. Subsurface storage and artificial recharge involve the concept that highly treated municipal and other wastewaters are valuable and should be reused. In a nation where water deficiencies or management problems are forecast for the foreseeable future, storage of treated wastewaters for reuse is destined to become a major element for consideration in water resource management (15).

Injection for the extraction of salt started in France in the 9th century and later in China. The first documented project for the disposal of oil field brine (saltwater produced along with oil and gas), in the same formation where it originated, started in Texas in 1938. Enhanced recovery of oil, which is the injection of water or other fluid into a formation to extract additional oil and gas, probably started early in the 1930s. Industrial waste injection started in 1950 with Dow Chemical injecting industrial fluids. In the 1950s DuPont Chemical Corporation started to inject some of its industrial waste into deep wells (16).

The earliest use of an injection well for municipal wastewater disposal in the US was in 1959 at the Collier Manor Sewage Treatment Plant in Pompano Beach, Florida (3, 13). During 13 years of operation, the City of Pompano Beach injected about three billion gallons (11.4 Mm^3) of secondary treated wastewater into a cavernous "boulder zone" through two wells 1000 to 1400 ft (305 to 427 m) deep (17).

Underground space is recognized (18) as a natural resource of considerable value. A small percentage of this space, like the "boulder zone," consists of large caverns capable of receiving and transmitting extremely large volumes of wastewater for a single injection well. But most space underground consists of the area available between sand grains in the rock strata. The percentage of this space available for fluid storage and movement depends upon how much clay and silt is present and the amount and type of cementing material present (19).

The porosity of a rock is essentially its interstitial pore space. It can be expressed quantitatively as the ratio of the volume of the pore space to the total volume of the rock, and generally is stated as a percentage. Gravel or sand that is clean, uniform and free of clay and silt will have about 30% to 40% of its volume available for storage space. Gravel or sand containing abundant clay and silt, cementing material, or a precipitant from injected wastes may have as little as 5% to 15% of its volume available. Fractures, joints, and solution channels in cemented rock formations, such as limestone, are additional types of pore space that contribute to porosity, but are difficult to measure.

Virtually all of this subsurface pore space is already occupied by natural water, either fresh or mineralized to some extent. Thus injection does not usually involve the filling of unoccupied space; but rather consists of the compression or displacement of existing fluids. Because the compressibility of water is small, creation of significant volumes of storage space through this mechanism requires disposal strata that underlie a large geographic area.

2. REGULATIONS FOR MANAGING INJECTION WELLS

The most accessible freshwater is stored in geological formations called aquifers. These aquifers feed our lakes; provide recharge to our streams and rivers, particularly during dry periods; and serve as resources for 92% of public water systems in the United States (2). Many people in the country also rely on ground water for their private drinking water wells. Injection of fluids can potentially contaminate aquifers that supply drinking water to households and public water systems. Direct injection into a drinking water aquifer, injection into a zone that is not isolated from a drinking water aquifer or poor performance of an injection well can contaminate ground water. Because contamination of ground water can be very persistent and difficult to remediate, it is important to ensure that contaminants do not enter ground water (16).

US EPA defines an injection well as any bored, drilled or a driven shaft or a dug hole, where the depth is greater than the largest surface dimension that is used to discharge fluids underground. This definition covers a wide variety of injection practices that range from more than 100,000 technically sophisticated and highly monitored wells which pump fluids into isolated formations up to 2 miles below the Earth's surface, to the far more numerous on-site drainage systems, such as septic systems, cesspools, and storm water wells, that discharge fluids a few feet underground (2).

Congress, in passing the Safe Drinking Water Act (SDWA) in 1974, gave US EPA the authority to control underground injection to protect underground drinking water sources (14). In 1979 and then again in 1980, US EPA developed the Statement of Basis and Purpose for the UIC (Underground Injection Control) Program, to support the regulations that were proposed

and then finalized in those years. US EPA published final technical regulations for the UIC program in 1980. The regulations set minimum standards state programs must meet to receive primary enforcement responsibility (primacy) of the UIC program, In 1981 Congress passed amendments to the SDWA that allowed for the delegation of the UIC program for injection wells to states if the program was effective in protecting underground sources of drinking water (USDW) and included traditional program components such as oversight, reporting and enforcement (16).

The US EPA groups underground injection into five classes for regulatory control purposes (2, 20). Each class includes wells with similar functions, and construction and operating features so that technical requirements can be applied consistently to the class. Class I includes the emplacement of hazardous and nonhazardous fluids (industrial and municipal wastes) into isolated formations beneath the lowermost USDW (3). Because they may inject hazardous waste, Class I wells are the most strictly regulated and are further regulated under the Resource, Conservation and Recovery Act. Class II includes injection of brines and other fluids associated with oil and gas production; Class III encompasses injection of fluids associated with solution mining of minerals; Class IV addresses injection of hazardous or radioactive wastes into or above a USDW and is banned unless authorized under other Statutes for ground water remediation. Class V includes all underground injection not included in Classes I to IV. Class V wells inject nonhazardous fluids into or above a USDW and are typically shallow, on-site disposal systems. Injection practices or wells that are not covered by the UIC Program include other individual residential waste disposal systems that inject only sanitary waste and commercial waste disposal systems that serve fewer than 20 persons that inject only sanitary waste.

Class I injection wells are sited such that they inject below the lowermost USDW and a confining zone above an injection zone (21). Injection zone reservoirs typically range in depth from 1700 to over 10,000 ft below the surface (3). Class I wells are mainly used in the following industries: petroleum refining, metal production, chemical production (22), pharmaceutical production, commercial disposal, municipal disposal, and food production There are 272 active Class I injection facilities nationwide. Of these, 51 are hazardous and 221 are non-hazardous. These 272 facilities maintain approximately 529 Class I injection wells that are scattered throughout the US in 19 states (3). The greatest concentrations are located in the Gulf Coast, Great Lakes, and the Floridian peninsular geographical regions.

The oil and gas injection wells, Class II account for a large proportion of the fluids injected in the subsurface (23, 24). Typically, when oil and gas are extracted, large amounts of saltwater (brine) are also brought to the surface. This saltwater can be very damaging if it is discharged in surface water. Instead, all states require that this brine be injected into formations similar those from which it was extracted (25). Over two billion gallons of brine are injected daily into injection wells in the US (26).

Mining wells, Class III are used in the mining of a number of minerals. In general the technology (27) involves the injection of a fluid, usually called lixiviant, which contacts an ore that contains minerals that dissolve in the fluid. The pregnant fluid (lixiviant nearly saturated with components of the ore) is pumped to the surface where the mineral is removed from the fluid. US EPA protects drinking water from contamination from mining wells by

implementing regulations that set minimum standards. These regulations require mining well operators to: case and cement their wells to prevent the migration of fluids into an underground drinking water source; never inject fluid between the outer-most casing and the well bore; and test the well casing for leaks at least once every 5 years.

Some of the practices using mining wells are (27): (a) Salt solution mining started in France in the 9th century. The Chinese used boring techniques to develop wells and extract brines at the beginning of the 18th century. In the US the salt industry started recovering brine at the end of the 18th century. The process consists of pumping water into the salt formation and extracting the salt from the resulting fluid after retrieval. More than 50% of the salt used in the US is obtained this way. (b) In-situ leaching of uranium is the practice of injecting a fluid to leach out the uranium salts and pumping it back to the surface where the uranium is extracted. 80% of the uranium extracted in the US is produced this way. (c) The production of sulfur using the Frasch process is one of the earliest uses of the technology. Traditionally, super heated steam is injected to recover a sulfur solution.

Shallow hazardous and radioactive injection wells, Class IV are prohibited unless the injection wells are used to inject contaminated ground water that has been treated and is being injected into the same formation from which it was drawn. These wells are authorized by rule (28) for the life of the well if such subsurface emplacement of fluids is approved by US EPA, or a State, pursuant to provisions for cleanup of releases under the Comprehensive Environmental Response, Compensation, and Liability Act of 1980 (CERCLA) or pursuant to requirements and provisions under the Resource Conservation and Recovery Act (RCRA).

The wells in the shallow injection wells, Class V are as diverse as they are similar. This category came about after all the easy definable wells were put into classes I through IV. In September 1999, US EPA completed The Class V notice of Underground Injection Control Study a comprehensive study proposed of most types of Class V wells (29). In May 2001, US EPA determined that existing federal underground injection control (UIC) regulations at the time are adequate to prevent Class V injection wells from endangering underground sources of drinking water (30). The final determination would be based on the Agency's evaluation of existing data collected for The Class V Underground Injection Control Study (EPA/816-R-99-014). In June 2003, Septic system and storm water discharges were regulated as Class V Wells (816-F-03-002 and 816-F-03-001 respectively).

Typically, Class V injection wells are shallow "wells," such as septic systems and drywells, used to place nonhazardous fluids directly below the land surface. However, Class V wells can be deep, highly sophisticated wells. The US EPA estimates there are more than 650,000 Class V wells in the United States (29). Class V wells are located in every state, especially in unsewered areas where the population is likely to depend on groundwater for its drinking water source.

Class V wells are convenient and inexpensive means to dispose of a variety of nonhazardous fluids. Some Class V wells are agricultural drainage wells, storm water drainage wells, large capacity septic systems, wastewater treatment effluent wells, spent brine return flow wells, mine backfill wells, aquaculture waste disposal wells, solution mining wells, in-situ fossil fuel recovery wells, special drainage wells, experimental wells, aquifer remediation wells, geothermal electric power wells, geothermal direct heat wells, heat pump/air conditioning

return flow wells, saltwater intrusion barrier wells, aquifer recharge and aquifer storage and recovery wells, subsidence control wells, and industrial wells (29).

Injection wells have the potential to inject contaminants that may cause our underground sources of drinking water to become contaminated. The UIC Program prevents this contamination by setting minimum requirements. The goals of the US EPA's UIC Program are to prevent contamination by keeping injected fluids within the well and the intended injection zone, or in the case of injection of fluids directly or indirectly into a USDW, to require that injected fluids not cause a public water system to violate drinking water standards or otherwise adversely affect public health. These minimum requirements affect the siting of an injection well, and the construction, operation, maintenance, monitoring, testing, and finally, the closure of the well (2). All injection wells require authorization under general rules or specific permits. Finally, States may apply to have primary enforcement responsibility (primacy) for the UIC Program. US EPA provides grant funds to all delegated programs to help pay for program costs. States must provide a 25% match on US EPA funds (31). To date (August, 2004), 33 States, Guam, the Commonwealth of the Mariana Islands and Puerto Rico have obtained primacy for all classes of injection wells. Seven States share primacy with the US EPA. The US EPA administers UIC programs for the remaining 10 States, the Virgin Islands, American Samoa and Indian Country (102).

3. BASIC WELL DESIGNS

Design of a casing program depends primarily on well depth, character of the rock sequence, fluid pressures, type of well completion, and the corrosiveness of the fluids that will contact the casing. Where fresh groundwater supplies are present, a casing string (surface casing) is usually installed to below the depth of the deepest groundwater aquifer immediately after drilling through the aquifer (Fig. 14.1). One or more smaller diameter casing strings are then set, with the bottom of the last string just above or through the injection horizon, the latter determination depending on whether the hole is to be completed as an open hole or gravel-packed, or is to be cased and perforated.

The annulus between the rock strata and the casing is filled with a cement grout. This is done to protect the casing from external corrosion, to increase casing strength, to prevent mixing of the waters contained in the aquifers behind the casing, and to forestall travel of the injected waste into aquifers other than the disposal horizon.

Cement should be placed behind the complete length of the surface casing and behind the entire length of the smaller diameter casing strings also, or at least for a sufficient length to provide the desired protection. It is suggested that at least $1\frac{1}{2}$ in. of annular space be allowed for proper cementing. Casing centralizers, other equipment, and techniques such as stage cementing can give added assurance of a good seal between the strata and the casing and should be encouraged where applicable.

The majority of injection wells constructed to date have been for one of three basic purposes: injection plus external monitoring for leakage through confining layers near the

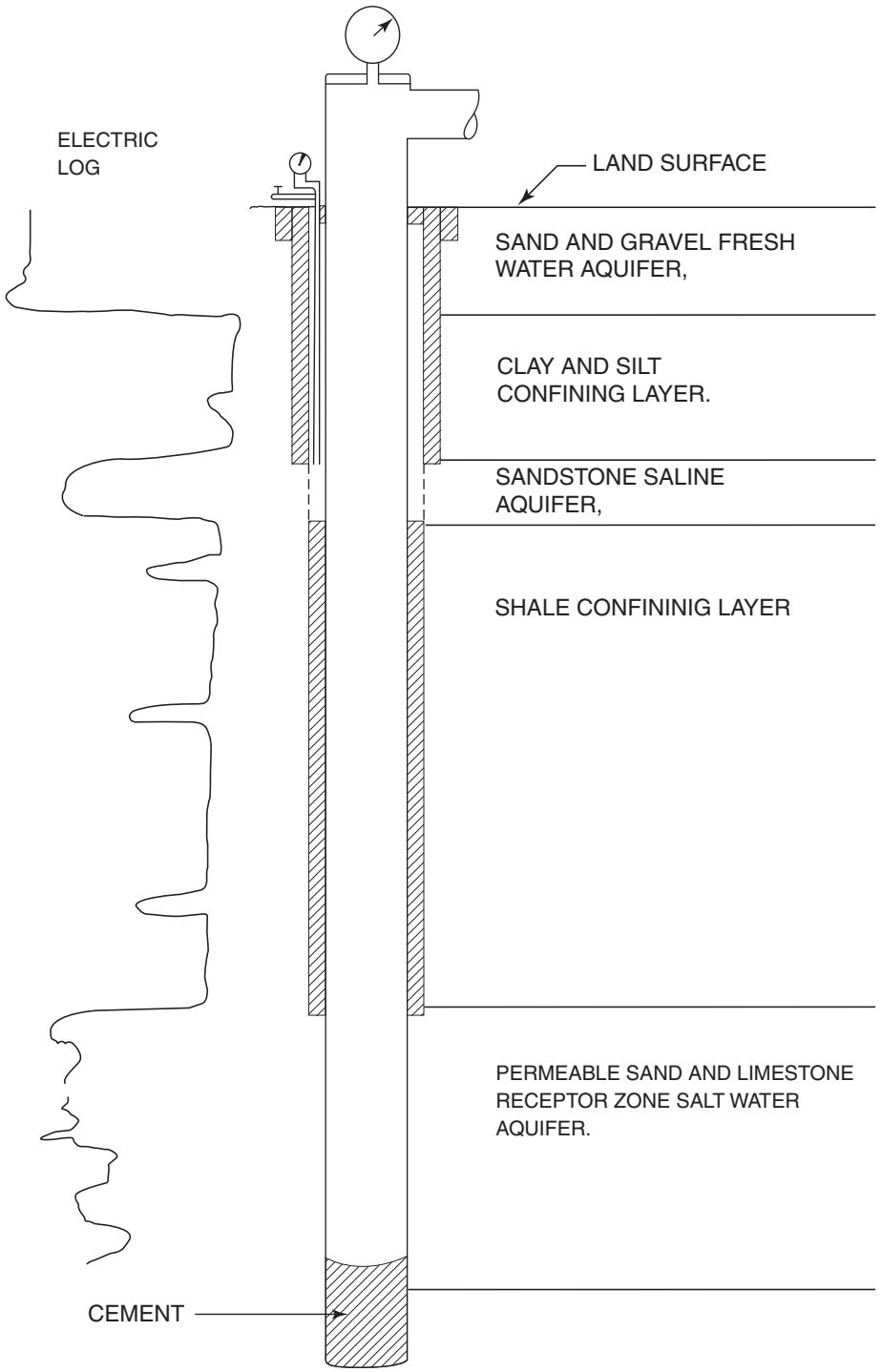


Fig. 14.1. Typical well construction. (*Source*: US EPA).

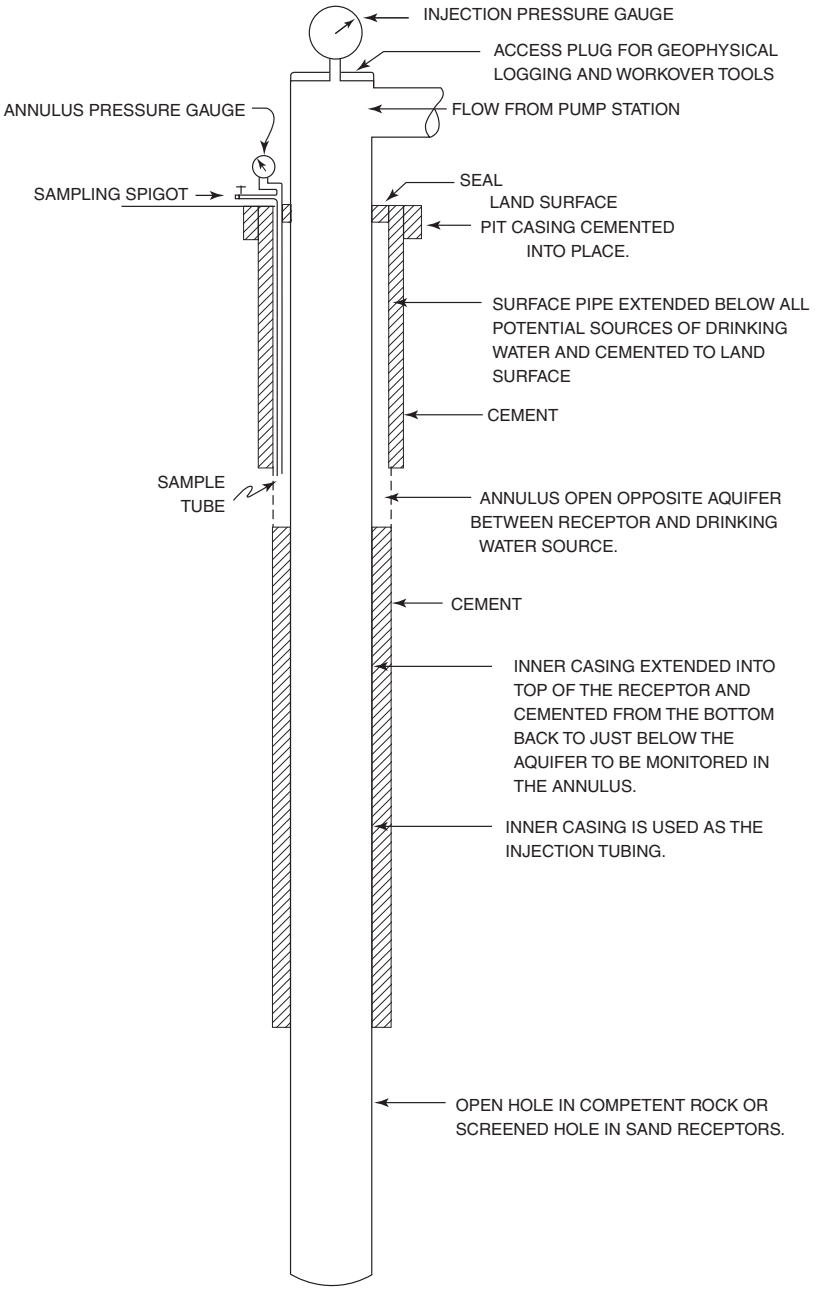


Fig. 14.2. Well designed for external self monitoring. (*Source*: US EPA).

well (Fig. 14.2); injection plus internal monitoring for leaks in the injection tubing and casing (Fig. 14.3); and injection only (Fig. 14.4). Numerous variations in design have been used to accomplish these three basic purposes.

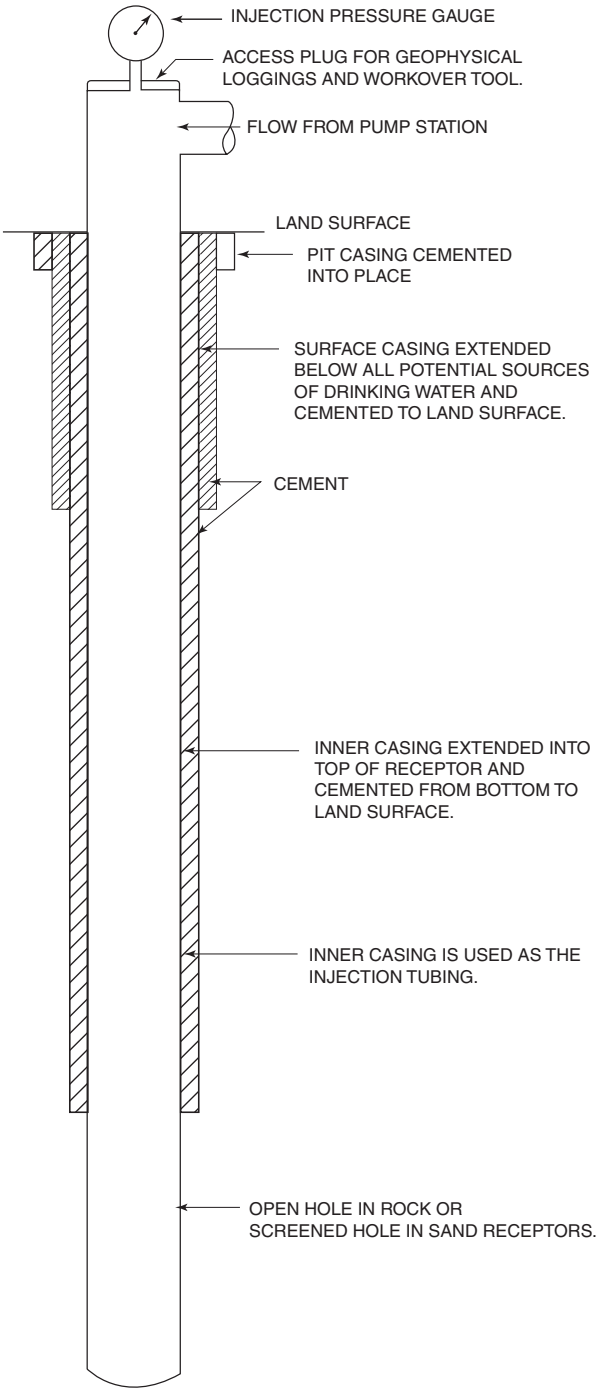


Fig. 14.3. Well designed for internal self monitoring. (*Source*: US EPA).

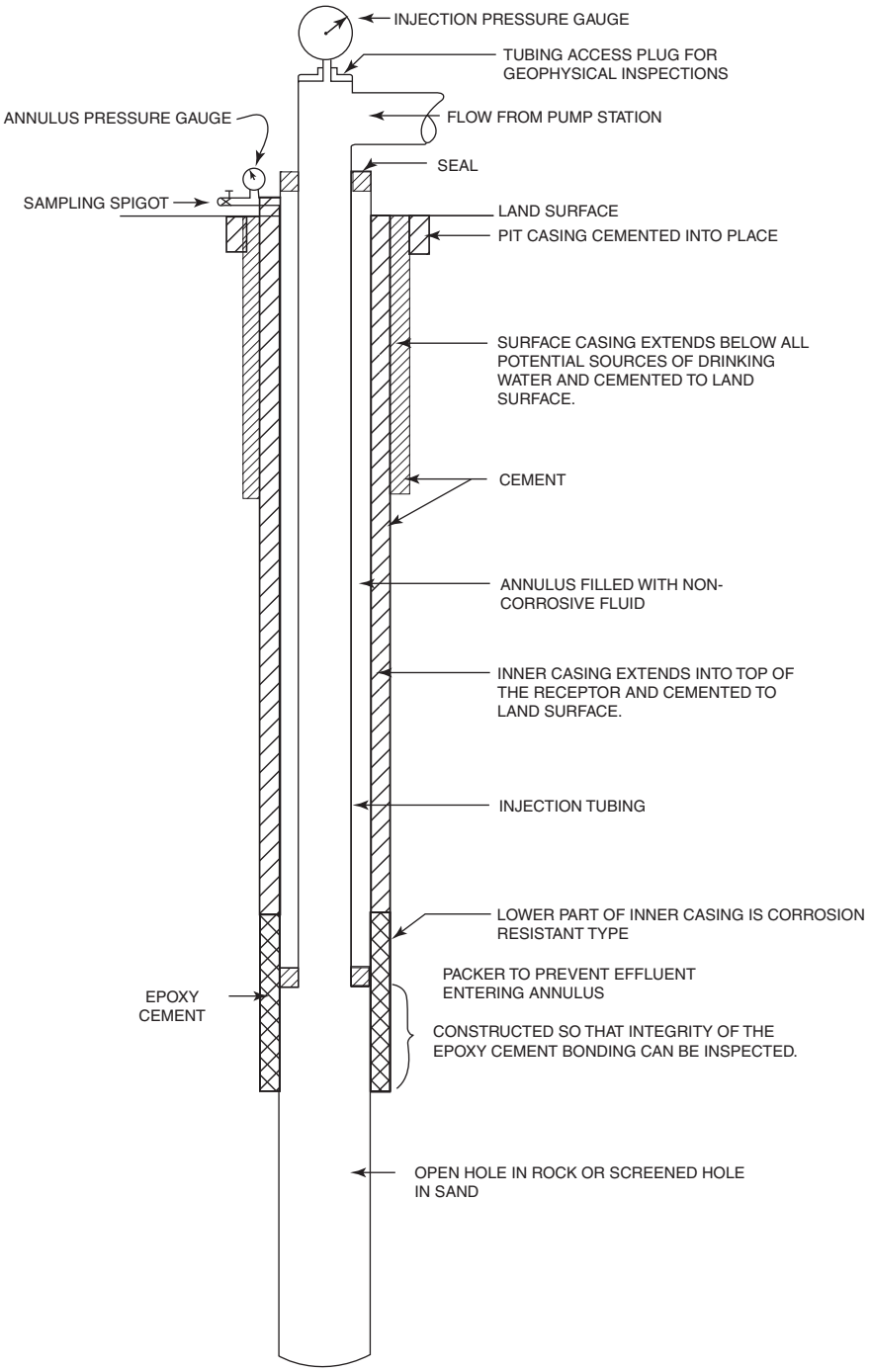


Fig. 14.4. Well designed for injection only. (*Source*: US EPA).

Wells designed for injection plus monitoring of the confining layer generally are constructed like the well in Fig. 14.2. In these type wells, the fluid chemistry and pressures outside the well in an aquifer overlying the receptor are monitored at the wellhead for changes that would indicate leakage. This monitoring is accomplished either by leaving the annulus exposed to the aquifer to be monitored as in Fig. 14.2, or by some method of accessing the aquifer inside the injection well casing. In designing a well for this purpose, the engineer should remember that to obtain a sample of fluids from the monitored aquifer, fluids in the pipes must be pumped out. If these fluids are saline, then their disposal may be a problem. One variation is to fasten a small 0.25 to 0.5 in. pipe (usually stainless steel or neoprene) to the outside of the inner casing (in the annulus) extending from land surface to the top of the aquifer being monitored. By sampling through this "drop" pipe, the volume of fluids to be disposed is minimized. Another variation is to drill a 4 to 6 in. larger hole and then attach a 2 to 3 in. monitoring pipe outside the inner casing (in the annulus) from land surface to the aquifer being monitored.

Wells designed for self-inspection of internal leaks are constructed similar to the well in the Fig. 14.3. A sampling tube is installed inside the casing to a depth near the bottom of the injection well. Either a seal is placed outside of this tubing near its bottom to prevent fluid circulation in the annulus and the annulus is pressurized, or low density fluid capable of causing back pressure, such as kerosene, is placed in the annulus. Pressure is measured at the land surface to detect fluid movement into or out of the annulus. Pressure changes in the annulus must be correlated with changes in injection rate, temperature, specific gravity of the fluids, and other factors to avoid false interpretation. In wells using this type of design, leaks in both the injection tubing and the well casing are quickly detected. Industrial waste should be injected through separate interior tubing rather than the well casing itself. This is particularly important when corrosive wastes are being injected. A packer can be set near the bottom of the tubing to prevent corrosive wastes from contacting the casing. Additional corrosion protection can be provided by filling the annular space between the casing and the tubing with oil or water containing a corrosion inhibitor.

Wells designed for injection purposes only are less expensive to construct, but lack the ability for self-monitoring described above. The well shown in Fig. 14.4 is typical of this type design. Separate wells must be drilled for monitoring purposes where required.

Temperature logs, cement bond logs, and other well-logging techniques can be required as a verification of the adequacy of the cementing. Cement should be pressure-tested to assure the adequacy of a seal.

Neat Portland cement (no sand or gravel) is the basic material for cementing. Many additives have been developed to impart some particular quality to the cement. Additives can, for example, be selected to give increased resistance to acid, sulfates, pressure, temperature, and so forth. Other additives reduce the viscosity of the cement until it flows like water.

It is frequently desired to increase the acceptance rate of injection wells by chemical or mechanical treatment of the injection zone. Careful attention should be given to stimulation techniques, such as hydraulic fracturing, perforating and acidizing to insure that only the desired intervals are treated and that no damage to the casing or cement occurs.

The type of well-head equipment can be a consideration in cases where the buildup of high back-pressure is a possibility. In such cases, the well head should be designed to "bleed-off" back flows into holding tanks or pits before pressures reach a hazardous level. High back-pressures can be developed by chemical reactions in the formation. For example, at Louisville, KY, where ferric chloride solutions had been injected into dolomite and limestone, for several years an excessive buildup of carbon dioxide gas pressure caused a blowout during routine maintenance in 1980.

Surface equipment often includes holding tanks and flow lines, filters, other treatment equipment, pumps, monitoring devices, and stand by facilities.

Surface equipment associated with an injection well should be compatible with the waste volume and physical and chemical properties of the waste to insure that the system will operate as efficiently and continuously as possible. Experience with injection systems has revealed the difficulties that may be encountered because of improperly selected filtration equipment and corrosion of injection pumps.

Surface equipment should include well-head pressure and volume monitoring equipment, preferably of the continuous recording type. Where injection tubing is used, it's advantageous to monitor the pressure of both the fluid in the tubing and in the annulus between the tubing and the casing. Pressure monitoring of the annulus is a means of detecting tubing or packer leaks. An automatic alarm system should signal the failure of any important component of the injection system. Filters should be equipped to indicate immediately the production of an effluent with too great an amount of suspended solids.

4. EVALUATION OF A PROPOSED INJECTION WELL SITE

It would be impossible to cover all the potential problems that could develop during the construction and operation of a disposal well system. But a safe economical injection well system that will function properly can be built at most sites in geographic areas suitable for this method of disposal.

Before proceeding with an evaluation, answers to the following two questions should be obtained from the regulatory authorities.

1. What criteria will determine the degree of pretreatment that will be required before fluids can be injected?
2. What restrictions have been placed upon water quality in receptor aquifers?

Pretreatment requirements vary from state to state. One state authority may prohibit emplacement of toxic wastes underground, whereas another will prohibit its surface discharge and encourage injection. At least one state agency pushes for pretreatment of all wastes to drinking water standards before injection. Many state agencies and most environmental groups push for "nondegradation" of the environment. But for successful operation of the system, all wastewaters must be pretreated until the fluid to be injected is compatible with the environment in the receptor aquifer.

A common mistake made evaluating deep-well injection systems is to underestimate the degree of pretreatment required. After a well system is completed is not the time to find that

after the extensive pretreatment required the wastes are in fact suitable for surface disposal. But before pretreatment can be determined, certain basic relationships between disposal operations and the physical environment must be examined. The important areas include: the receptor zones, the confinement conditions and the subsurface hydrodynamics. Data on these subjects are available from state and federal agencies, as well as local consultants, and should be assembled and evaluated. If the results are favorable, then additional data should be collected by constructing test wells. Various potential confining layers and receptors then need to be tested. Final evaluation of data may show the disposal of wastewaters into underground formations to be an unwise solution to a disposal problem not only from the standpoint of damage to the environment, but from the standpoint of overall costs.

4.1. Confinement Conditions

Confining layers, although generally required by regulatory agencies, are not always essential. For example, in one of the southeastern states, wastes are being injected into the Knox Dolomite where it is about 5000 ft (1500 m) thick and intensely fractured vertically. Porosity is low, averaging about 10%. Wastes being injected are heavy, having a specific gravity of about 1.2. The heavy wastes after injection move in response to gravity downward to the base of the Knox where they are permanently stored. The saltwater-freshwater boundary that now lies at about 4000 ft (1200 m) is being displaced upward at a rate of about 1 ft/year.

Confining layers are rarely impermeable to waste movements. They just retard the movement. Most wastes are capable of slowly moving through even the denser clays. However, as movement takes place, the wastes are subjected to ion exchange, osmosis, filtration, absorption and other forms of treatment. The rate of leakage through a confining bed (Q) and the velocity of fluid movement (v) can be determined from Eqs. (1) and (2).

$$\mathrm{Q} = PIA \tag{1}$$

$$v = PI/\Phi \tag{2}$$

where:

Q = rate of leakage, ft^3/day
v = velocity of fluid, ft/day
A = leakage Area, ft^2
P = permeability, $ft^3/day/ft^2$
I = hydraulic gradient, ft/ft
ϕ = porosity expressed as a decimal

Possibly the most important laboratory tests to be conducted on samples collected during drilling of the test wells are vertical permeability and ion exchange capacity on cores collected from the confining layer overlying the receptor. The vertical permeability, together with thickness of the confining bed and the anticipated differential pressure across the confining bed, should be used to predict the velocity at which wastewater will travel through the

Table 14.1
Relationship of the coefficient of permeability to potential of a stratum for use as a receptor or as a confining layer

Rock type	Flow potential	Permeability range, m/d
Cavernous limestone	Excellent receptor	3×10^6 to 1×10^9
Gravel	Good receptor	1×10^4 to 3×10^6
Sands and sandy silts	Poor receptor or confining layer	1×10^{-2} to 1×10^4
Clay, shale	Good confining layer	1×10^{-6} to 1×10^{-2}

confining bed. The general permeabilities of rocks are shown in Table 14.1. Data generated from ion exchange and other tests may show that toxic wastes can meet drinking water use or other applicable standards after being subjected to subsurface treatment. These tests also indicate the physical and chemical changes that may take place with time during movement through the confining bed.

4.2. Potential Receptor Zones

A basic requirement of a receptor zone or a combination of zones is that it be capable of receiving and transmitting the volume of wastewater planned for injection. State regulatory agencies frequently place additional requirements on receptors based upon depth and water chemistry. For example, some state agencies prohibit injection into receptors at a depth less than 2000 ft (610 m) or where the native fluids contain less than 10,000 mg/L total dissolved solids.

Another requirement is that changes in the physical and chemical properties of the wastewater and in the receptor after injection be compatible with the goals of injection. These changes usually can be grouped as follows:

1. Changes in the wastewater induced by the environment in the receptor
2. Changes in the wastewater caused by chemical reactions with the receptor rocks
3. Changes in the wastewaters caused by chemical reactions with fluids in the receptor and
4. Physical and chemical changes in the receptor resulting from reactions with the wastewater

Precipitants formed as a result of these types of changes can plug the receptor and cause the system to fail.

Knowledge of the complete chemical character of the wastewater after the pretreatment is of the foremost importance in evaluating a potential receptor. This knowledge, plus data about the physical subsurface environment available from drilling oil wells in the area or from a test well drilled at the proposed site, should enable a company to forecast the chemical stability of its waste. Knowledge of the mineralogy of the aquifer and the chemistry of interstitial fluids collected from a test well should indicate the reactions to be anticipated during injection. Laboratory tests can be performed with rock cores, formation fluids, and wastewater samples to confirm anticipated reactions.

Selm and Hulse (32) list reactions between injected and interstitial fluids that can cause the formation of plugging precipitates. These include precipitation of alkaline earth metals such as calcium, barium, strontium and magnesium as relatively insoluble carbonates, sulfates, orthophosphates, fluorides, and hydroxides. Precipitation of other metals such as iron, aluminum, cadmium, zinc, manganese, and chromium as insoluble carbonates, bicarbonates, hydroxides, orthophosphates, and sulfides can also occur. Also, the precipitation of oxidation-reduction reaction products can occur.

Carbonate rocks generally are excellent receptors for acid wastewaters. The soluble carbonate rocks neutralize the acid and cause precipitation of many of the above metals. Where the volume of precipitant is significantly less than the volume of carbonate rock dissolved, the system will work safely. If not, then the receptor pores will generally plug and the system will fail. Undesirable effects of the reaction of acid wastes with carbonate receptors could be the evolution of carbon dioxide gas that might retard fluid movement if present in excess of its solubility.

Marine sand receptors containing clays such as montmorillonite will pass saline wastewater without change, but the clays may swell to many times their original volume when in contact with freshwater. Such swelling effectively reduces permeability and may cause well failures.

4.3. Subsurface Hydrodynamics

The dynamics of subsurface fluids in the receptor and overlying aquifers must be understood to the extent that the direction and rate of movement of any wastewater injected and any native fluids displaced by this waste can be estimated. Using data collected from a test well and other means, the pressure buildup in affected aquifers with time should be estimated.

A well injecting at a constant rate into an extensive confined receptor aquifer produces an area of influence that expands with time. As the formation pressure is increased, flow is radially away from the injection well, but not a steady-state flow. Theis (33) hypothesized a close analogy between groundwater flow and heat conduction, and developed the following nonequilibrium equation (Fig. 14.5) for determination of the coefficients of transmissibility (T) and storage (S).

$$T = \frac{114.60QW(u)}{s} \tag{3}$$

$$S = uT/1.87(r^2/t) \tag{4}$$

where:

T = coefficient of transmissibility
S = coefficient of storage
Q = flow or injection rate (gpm or barrels/day)
$W(u)$ = the well function of u as shown in Table 14.2,
s = the pressure change at r, ft of water
r = the radius from the point of injection to the point of observation, ft
t = the time of injection, day

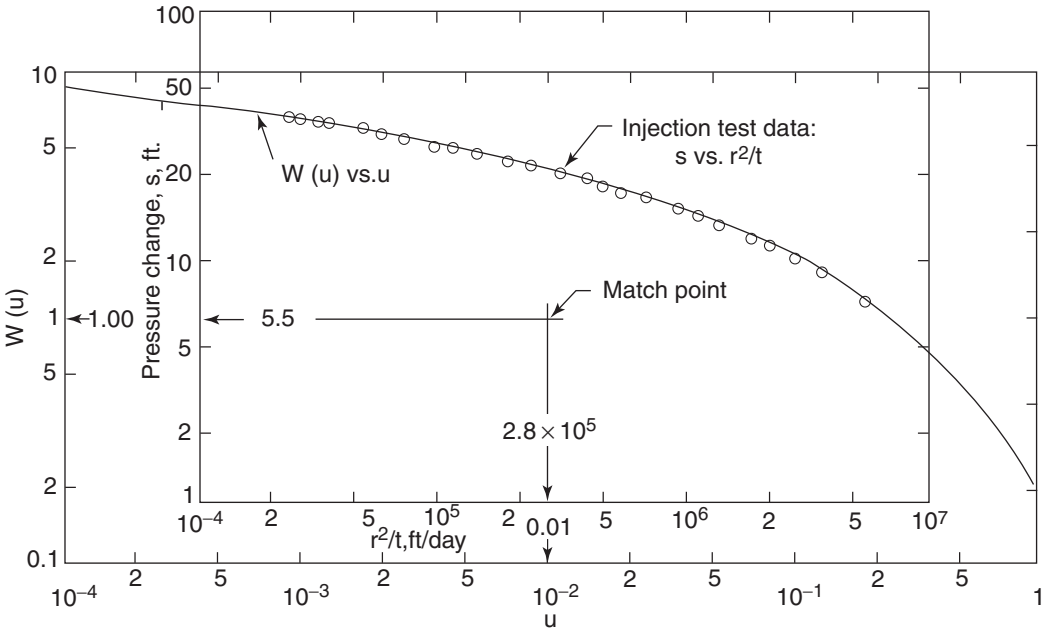


Fig. 14.5. Theis method of superposition for solution of the nonequilibrium equation. (*Source*: US EPA).

Table 14.2
Values for $W(u)$ for values of u

u	1.0	2.0	3.0	4.0	5.0	6.0	7.0	8.0	9.0
$\times 10^{-1}$	1.82	1.22	0.91	0.70	0.56	0.45	0.37	0.31	0.26
$\times 10^{-2}$	4.04	3.35	2.96	2.68	2.47	2.30	2.15	2.03	1.92
$\times 10^{-3}$	6.33	5.64	5.23	4.95	4.73	4.54	4.39	4.26	4.14
$\times 10^{-4}$	8.03	7.94	7.53	7.25	7.02	6.84	6.69	6.55	6.44
$\times 10^{-5}$	10.94	10.24	9.84	9.55	9.33	9.14	8.99	8.86	8.74
$\times 10^{-6}$	13.24	12.55	12.14	11.85	11.63	11.45	11.29	11.16	11.04
$\times 10^{-7}$	15.54	14.85	14.44	14.15	13.93	13.75	13.60	13.46	13.34
$\times 10^{-8}$	17.84	17.15	16.74	16.46	16.23	16.05	15.90	15.76	15.65
$\times 10^{-9}$	20.15	19.45	19.05	18.76	18.54	18.35	18.20	18.07	17.95
$\times 10^{-10}$	22.45	21.76	21.35	21.06	20.84	20.66	20.50	20.37	20.25
$\times 10^{-11}$	24.75	24.06	23.65	23.36	23.14	22.96	22.81	22.67	22.55
$\times 10^{-12}$	27.05	26.36	25.96	25.67	25.44	25.26	25.11	24.97	24.86
$\times 10^{-13}$	29.36	28.66	28.26	27.97	27.75	27.56	27.41	27.28	27.16
$\times 10^{-14}$	31.66	30.97	30.56	30.27	30.05	29.87	29.71	29.58	29.46
$\times 10^{-15}$	33.96	33.27	32.86	32.58	32.35	32.17	32.02	31.88	31.76

Source: Wenzel (43).

The Theis method is a graphical one based on superposition of curves. $W(u)$ is plotted against u on logarithmic paper then s is plotted against r^2/t using paper of the same scale. The two plots are superimposed with the coordinate axes parallel and shifted until the position with most of the two curves matched is found as shown in Fig. 14.5. The coincident values of $W(u)$, s and r^2/t are noted. By substituting these values into the above formulas, S and T for the receptor aquifer are determined. See example 3 in Section 11.

5. POTENTIAL HAZARDS-WAYS TO PREVENT, DETECT, AND CORRECT THEM

Problems with injection wells generally can be related to failures in one or more of the five areas listed below (34, 35):

1. Lack of consideration of all fluid movements
2. Failure of the receptor to receive and transmit the wastes
3. Failure of the confining layer
4. Failure of an individual well either in design, construction or operation
5. Failure because of human error

Experience and the use of good common sense will avoid most of these problems. When injection wells arc properly designed, installed, operated, and maintained, they are no different from any other good piece of equipment. If properly used, they can play an important role in removing toxic and hazardous substances from the immediate human environment.

5.1. Fluid Movement during Construction, Testing, and Operation of the System

Because of lack of knowledge about fluid movements, one drilling company placed saltwater generated during pump testing of a receptor Zone into an unlined storage pit on their site. The saltwater, because of its heavy specific gravity seeped downward into the local drinking water aquifer. It then moved laterally without mixing and some 8 months later contaminated a nearby city's drinking water supply.

Each well should be designed so that local freshwater supplies are protected from contamination by a separate casing set into the top of their underlying confining layer and cemented back to land surface before the confining layer is breached during construction. Mud pits should be lined with impervious material to prevent seepage into the shallow freshwater strata. All material used during construction, such as salt, mud, acid, and the like, should be stored in such a way as to assure that materials spilled from damaged containers and the like will not contaminate the freshwater supply.

Another example of a problem is the company that sought to cut costs by eliminating consulting fees. They hired a local, very capable, water-well contractor to design and construct their injection and monitoring wells. He did an excellent job of designing the injection well, four monitoring wells into the receptor aquifer, plus one monitoring well into an overlying aquifer. The only problem was that he did not understand subsurface hydrodynamics and located these entire monitor wells within 150 ft of the injection well. The waste front passed the deep monitor wells within 3 days after injection started. Four replacement monitor wells

had to be drilled some distance away. Monies wasted on the first four wells (about $140,000) far exceeded the cost of hiring an experienced ground water geologist.

Making sure that the flow capacity in the test equipment is representative of the proposed permanent facilities under design is also important. An abandoned gas well being tested for conversion into a disposal well showed a capacity of only 50 gpm on gravity flow (36). The project was almost abandoned when an engineer realized that the friction loss in the temporary piping was higher than in the permanent facilities. After the piping was changed, the injection tests showed the well would handle 450 gpm on gravity flow.

5.2. Failure of the Aquifer to Receive and Transmit the Injected Fluids

Failure of an aquifer generally is caused by lack of naturally developed permeability in the receiving zone or by filling of the pore space with either suspended solids from the effluent or precipitants formed by chemical or biological reactions in the receptor. Well-head injection pressure should be continuously monitored so that these problems can be detected early and failure avoided. A reduction of the ability of the receiving zone to accept and transmit wastes from any cause will increase the pressure at the well head.

Plugging of the receptor zone is by far the most common operational problem where the receptor is a sand aquifer. Most plugging problems can be avoided by one or more of the following:

1. Detailed coring to study the size and shape of pore spaces in the receptor
2. Detailed chemical analyses of fluids and rocks in the receptor
3. Biological cultures of both receptor fluids and wastes
4. Analysis of pressures and temperatures in the receptor
5. Changes expected in the wastes after injection and
6. Proper cleanup during completion of the well

One injector ran extensive compatibility tests in his own labs at room temperatures and pressures with no indication that a problem might exist. However, a few weeks after injection began the injection pressures started rising above the predicted pressures. New tests conducted under environmental conditions similar to those in the receptor showed that minor changes in fluid pH caused precipitants to form that partially plugged the receptor.

Another injector's effluent was found to be incompatible with the native fluids in the receptor. If the pH was raised slightly a precipitant formed; when the pH was lowered, gases were released. The solution: keep the two fluids separated by injecting a compatible buffer ahead of the effluent. The importance of compatibility tests cannot be overemphasized. These tests should be run as close to actual well conditions as possible.

5.3. Failure of the Confining Layer

The effectiveness of the confining layer at each site should be thoroughly investigated by monitoring changes in formation pressures and chemical water qualities during testing of the disposal wells. The most frequent cause for such failures is the creation or the opening up of vertical fractures in a receptor (Fig. 14.6), then propagation of these fractures by continued

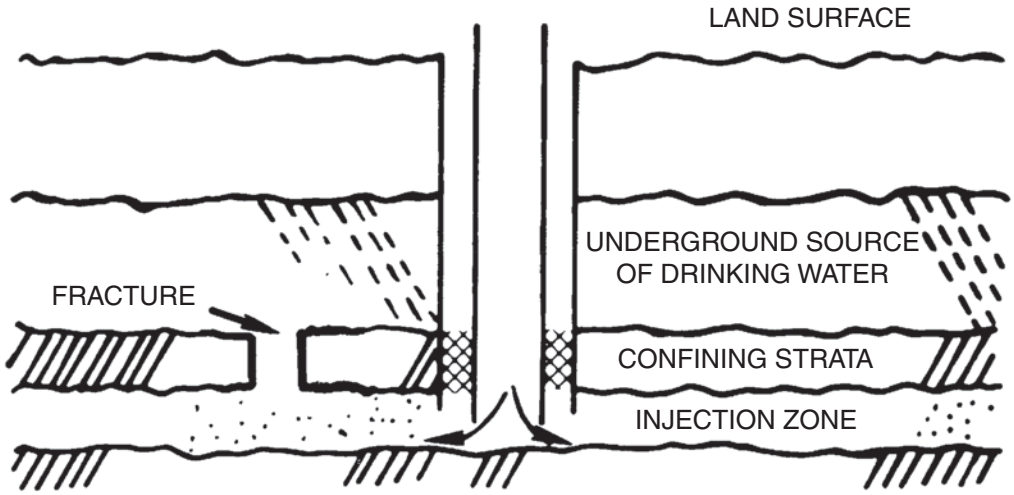


Fig. 14.6. Faulty or fractured confining strata. (*Source*: US EPA).

injection until they radiate through the confining layer. For example, a company was permitted to inject its wastes at a maximum surface pressure of 150 psi, a rather conservative pressure for most operations. But the specific gravity of the fluids was not taken into account by either the permitting agency or the company. During the operative history of this system, each time the injection pressure approached 149 psi it would miraculously (so the company officials thought) decline. Soon the pressure in nearby shallow observation wells began to rise. Then wastes were detected in these observation wells and the company had to abandon its injection well system. They built surface treatment facilities at great expense. The cause of the problem was uncontrolled vertical hydraulic fracturing of the confining layers at a pressure of 149 psi. The bottom-hole pressure was about 650 psi, well within the vertical fracturing gradient for the shallow unconsolidated silts and clays that made up the confining layer at this site.

Hydraulic fracturing can be avoided by adequately testing the injection system before injecting any effluent. Testing and operations should be planned so that operating pressures never exceed the bottom-hole pressures reached during testing.

It is almost inconceivable that an entire system would fail considering the operational history of existing wells and provided that all the precautions and testing described herein are incorporated into the evaluation of a site. However, should this occur, the disposal of effluent into the subsurface is riot a strictly one-way process. If pressures begin to build up or saline water begins to increase in an aquifer above the injection zone, which would indicate leakage of a saline front moving out ahead of the waste front, the entire disposal system could be reversed by installing pimps into the disposal wells and pumping the injected fluids back up to land surface. The effluent can then be disposed via an alternate method. Much of the

injected effluent can be recovered by this means. The remaining effluent would probably stop moving as pressures are reduced.

5.4. Failure of an Individual Well

An example of extreme well failure was the injector who went to great expense to install and cement into place fiberglass casing through which they could inject acids without installing injection tubing. However, they overlooked the fact that a cement plug had to be drilled out of the bottom of the fiberglass casing, the hole deepened and screens installed. Of course the driller did not use centralizers and his drill stem fractured the fiberglass. With time acids in the effluent ate through the cement then entered a shallow aquifer above the monitoring point. This company now has abandoned its injection well system and has drilled more than 20 monitoring wells to keep track of the movement of wastes lost through the fiberglass casing. The local regulatory agency has declared that injection wells will never be used in their state again. Inspection of this well with a caliper upon completion would have detected the fractures, which would then have been repaired and the failure avoided.

After completion of all construction, each well should be thoroughly inspected and tested using all practical geophysical and hydrogeologic methods available (37). Constructing each well using the best available technology followed by extensive testing and inspection before its use for disposal of effluents is the best practical way to avoid well failure.

Should effluents leak through a break in the casing, packers can then be installed to isolate the break. The zone between the packers can be pumped to recover the leaked fluids. Leaks of this type can usually be repaired. It is recommended that an emergency standby well be constructed at each site so that in the event of mechanical failure of some of the pumping equipment or a need develops for servicing or inspecting an injection well, the well that needs servicing can be shut down, and the emergency standby well can be put into operation. Also this emergency standby well can be utilized for observational purposes to gain additional information on changes in water quality in the aquifer during injection.

5.5. Failures Because of Human Error

An example of human error is an injector who noticed a sharp increase in the injection pressure. The pressure continued to rise until the maximum head pressure of the pump was reached and then the flow began to decrease. Investigation of the problem revealed that one of the cartridges in the polishing filter had been left out.

Davis and Funk (36) cited another excellent example of the importance of human error at a site where a new disposal well system was installed and shut-in pending the startup of a new plant. Compatibility tests had shown the presence of freshwater-sensitive clays in the receptor. During plant construction the transfer lines to the disposal well systems were pressure-tested with freshwater by the contractor and left full. When the well was put into service by the company the freshwater was displaced into and plugged the receptor aquifer by causing hydration and swelling of the freshwater-sensitive clays. The result was a 6000 ft deep $250,000 posthole.

Experience is the best solution to human error. However, installing automatic shutoff and alarm systems at key locations in the system will minimize the effects of human error.

Table 14.3
Capital costs for injection well systems

Average Flow Q MGD	Cost 2007 Million US $
0.5	0.63
1.0	0.68
2.0	0.73
3.0	0.91
5.0	1.27
10	1.82
20	3.09
30	4.73
50	7.63
100	18.18

6. ECONOMIC EVALUATION OF A PROPOSED INJECTION WELL SYSTEM

Costs for constructing and operating an injection well system vary tremendously from one area of the country to another (38). They are lower in areas with active petroleum exploration because of competitive bidding, equipment availability, and the availability of energy sources. Costs shown herein are adjusted using US Army Corps of Engineers Cost Index for Utilities of 302.25 in 1981 to 539.74 in 2007 (39, Appendix). Costs are broken down into general costs and indirect costs.

Table 14.3 shows the range in capital costs experienced in the southeastern United States by industries and municipalities for the construction of injection well systems. An engineer should be able to take these cost data, adjust them for inflation and other differences in the local area, add pretreatment capital costs, and make a reasonable estimate of the total costs for disposal of a company's wastes by well injection. These costs are based upon an average well depth of 3500 ft (1067 m).

Operation and maintenance costs range from 80 cents to about $2.18/1000 gal waste water injected (2007 energy costs). This variation is caused primarily by differences in well-head injection pressures monitoring requirements. The average cost is about $1.42/1000 gal injected.

Legal fees, permitting fees, changes in insurance rates, and other indirect costs can amount to as much as 10% of construction costs and should be recognized during cost estimating.

7. USE OF INJECTION WELLS IN WASTEWATER MANAGEMENT

Numerous pro and con considerations must be investigated before deciding upon a method of discharge for a waste fluid. For industries in certain locations, the disposal or storage of wastes in the subsurface by use of well injection may be the most environmentally acceptable

practice available. The difference between storage and disposal is that storage implies the existence of a plan for recovery of the material within a reasonable time. Whereas disposal implies that no recovery of the material is planned at a given site. Either operation will require essentially the same type of information before injection. However, the attitude of the regulatory agencies toward evaluation of a proposal to use the well injection method could be quite different for each (1, 3, 7, 16, 21, 29).

7.1. Reuse for Engineering Purposes

The reinjection of fluids produced with hydrocarbons into oil-producing horizons to maintain oil field pressures has been practiced for years. Land subsidence in the vicinity of many oil and gas fields and in areas of substantial groundwater production is generally considered to be caused by the withdrawal of fluids. In California, for example, treated wastewaters are being injected to retard the rates of subsidence by repressurizing subsurface formations.

Along many coastal areas (40) the heavy withdrawal of potable ground water for municipal, industrial and other uses from freshwater aquifers has caused saltwater encroachment laterally within the aquifer systems. In such areas, treated wastewaters may be injected into the aquifer system to create a hydraulic barrier and hold back the encroaching saltwater (33, 41–43). For example, since 1965 tertiary-treated wastewater has been injected at Bay Park NY, into a shallow artesian sand aquifer used for public water supply (42) to create a hydraulic barrier against saltwater encroachment.

In areas with water shortages, or where for environmental reasons other methods of discharge are not practical, it may be desirable to inject highly treated wastewaters into the subsurface for purposes of ground water storage. Wells designed for recharge purposes will see increased use as a tool for capturing and storing highly treated wastewaters such as would come from drain tiles underlying industrial or municipal land spreading disposal sites. This method should be analyzed considering engineering feasibility and health effects.

7.2. Injection Wells as a Part of the Treatment System

The use of injection wells for subsurface treatment of the wastewaters deserves greater consideration. The emplacement of acid wastewaters in limestone, marble, dolomite, or other carbonate formations where the acid will be neutralized through chemical reaction with the rock formation is generally more environmentally sound than surface treatment with a combination of strip mining of carbonate rocks, plus surface discharge of neutralized wastewater, plus land tilling of the solids precipitated during neutralization (44, 45). A key requirement for successful well injection is the determination that the volume of solids that may be precipitated is substantially less than the volume of rock dissolved during neutralization. An example of this operation is Kaiser's disposal well system at Mulberry, Florida, where about 150 gal/min (9.4 L/s) of fluosilicic acid (pH less than 1.0) and sodium chloride (6.5%) is being injected into a porous dolomite (Lawson Formation of Upper Cretaceous age) at a depth of about 4500 ft (1372 m). The ratio of volume of dolomite dissolved to volume of precipitant formed is about 11:7. The permeability of the receiving aquifer is increasing with time, indicating that the system is operating as planned and that the precipitant is not plugging the formation pore space.

Where odors may be a problem, where the rate of treatment is slow or where economics are favorable, the subsurface can be used for chemical treatment and for biological treatment by anaerobic and facultative microorganisms. One of the world's largest nylon plants, Monsanto's Plant located near Pensacola, Florida, since 1963 has been disposing of its wastewaters containing nitric acid (pH 2.3) and other nitrogen compounds into the Ocala Limestone of the Eocene age. Back flushing experiments carried out by the US Geological Survey (45) in 1968 showed that the pH of the injected fluid increases rapidly in the aquifer accompanied by rapid denitrification and generation of carbon dioxide, nitrogen, methane, and other gases. Nitrate concentrations decreased from 3,000 to near 0 mg/L in less than 75 minutes.

Still another use of the subsurface for treatment is the storage of wastewaters containing radioactive minerals that have relatively short half-lives.

7.3. Storage of Municipal Wastewaters for Reuse

Under certain conditions, a double benefit can be realized by injecting a good quality wastewater effluent into a saline aquifer: potentially harmful viruses and bacteria that might survive the treatment process are removed from the human environment, and the effluent displaces a poorer quality (salty) groundwater, thus creating a reserve of potentially usable water in underground storage.

Several deep injection wells have been constructed in Florida, Hawaii, Louisiana, Illinois, and Texas for storage of secondary-treated wastewater effluent into saltwater aquifers. Secondary- and tertiary-treated municipal wastewaters are of such good quality and in such large volumes (5 to 50 MGD) (220 to 2,200 L/s) that it is much too valuable to waste in areas where water shortages are forecast for the foreseeable future. The storage of treated wastewater for future reuse is receiving increased attention in long range management planning (6, 11–13, 15, 40, 42, 46). Expansion of this method of reuse as a tool of long-range water quality and water resource management is being encouraged by the US Environmental Protection Agency (US EPA) (47) and many state regulatory agencies as long as measures are taken to protect the public health. The method is particularly adaptable and acceptable when the planned reuse is for agricultural or other nonpotable demands.

Esmail and Kimbler (48) in their investigation of the technical feasibility of storing freshwater in saline aquifers concluded that the rate of injection and the permeability of the receiving aquifer were the two principal factors that control the recovery of the stored freshwater. Recovery is inversely proportional to the aquifer's permeability and directly proportional to the rate of injection.

7.4. Storage of Industrial Wastewaters

In this era of rapidly changing economics, developing technology, increasing energy costs, and demands for reuse and recycling, what is today a waste product may tomorrow become a valuable byproduct (7). At each plant serious consideration should be given to separating streams of wastewater that contain chemicals with a potential for future reuse, these reusable chemicals could then be injected in a manner whereby they can be recovered at a future date,

7.5. Disposal of Municipal and Industrial Sludges

One of the developments is a method (49) of dewatering, compacting, and disposing of sludges and other solid wastes from municipal and industrial waste treatment plants using the elastic rebound properties of subsurface strata to compress and dewater the waste. Hydraulic pressure is used to compress the rock and create a large opening into which the sludge is placed.

Water separated from the sludge migrates through the receptor stratum radially away from the injection well, which is the point of greatest pressure. Immediately following injection, the volume of the opening in the receptor stratum is slightly less than, but directly proportional to, the volume of waste injected. But as dewatering takes place, the volume of the opening is reduced and sludge compaction begins. During compaction the volume reduction is proportional to the amount of its suspended solids content, for example, if the sludge contained 95% water and 5% solids then the compacted volume will be about one twentieth of the injected volume, Deformation properties such as elasticity and plasticity allow the overburden to absorb the increased thickness, except that if large volumes are emplaced at shallow depth a small but measurable rise in land surface would be expected.

The end product of this method is a series of very thin (0.001 to 0.01 in.) pancakes of sludge.

8. USE OF INJECTION WELLS FOR HAZARDOUS WASTES MANAGEMENT

Injection of hazardous wastes into deep wells began in the United States in the 1960s (3). At that time, the chemical industry was looking for a safe, relatively inexpensive method for disposing of high volumes of waste that could be considered toxic. Technology was borrowed from the oil and gas industry to develop this new form of disposal.

Currently (2002) there are 163 hazardous waste deep injection wells in the US located at 51 facilities. Most are found in Texas (78) and Louisiana (18). Eleven of the facilities are commercial hazardous waste injection facilities. These are the only facilities that can accept hazardous waste generated offsite for injection. Ten of them are located in the Gulf Coast region whereas one is located in the Great Lakes region (3).

A few regulatory agencies prohibit the disposal of toxic waste underground. Others require the best available pretreatment before injection on the premise that the safety of the method is maximized by this approach. The US Public Heath Service (50) considers that only concentrated toxic wastes that cannot otherwise be satisfactorily disposed of should be considered for deep-well injection. The US EPA (18) considered such disposal to be temporary until new technology becomes available, enabling more assured environmental protection. However, as the result of the Hazardous and Solid Waste Amendments of 1984, EPA published special regulations for deep wells injecting hazardous waste. In addition to making the requirements for these wells more stringent, the regulations require that each well operator provide a demonstration that the hazardous waste will not be released from the injection zone for at least 10,000 years, or will be rendered nonhazardous by natural processes (16).

The injection of troublesome industrial waste waters into subsurface formations via deep wells is a relatively simple low cost disposal procedure that has attracted the attention of many manufacturing companies, particularly of the refining and chemical industries (51–55).

Deep-well disposal of hazardous wastes has been demonstrated to be technically feasible in many areas of the country (5–9, 50). However, ill sited and improperly designed or constructed wells can result in serious pollution problems. The effects of subsurface injection and the fate of injected wastewaters should be adequately researched to ensure protection of the integrity of the subsurface environment.

8.1. Identification of Hazardous Wastes

Wastes are defined as hazardous for purposes of regulatory control in 40 CFR Part 261. In this regulation, wastes are classified as hazardous either by being listed in tables within the regulation or by meeting certain specified characteristics. Thus under 40 CFR Part 261 hazardous wastes are known either as *listed* or *characteristic wastes*. Some listed waste streams, such as spent halogenated solvents, come from many industries and processes. Other listed waste streams, such as API separator sludges from the petroleum- refining industry, come from one particular industry and one process. Radioactive wastes are not covered by 40 CFR Part 261. A characteristic waste is not listed, but is classified as hazardous because it exhibits one or more of the following four characteristics (56):

(a) ***Toxicity***

A waste is toxic if the extract from a representative sample of the waste exceeds specified limits for eight elements and four pesticides using extraction procedure (EP) toxicity test methods. The elements are arsenic, barium, cadmium, chromium, lead, mercury, selenium, and silver. The pesticides include endrin, methoxychlor, toxaphene, 2,4-D and 2,4,5-TP Silvex. Note that this narrow definition of toxicity relates to whether a waste is defined as hazardous for regulatory purposes; in the context of this chapter, toxicity has a broader meaning because many deep-well-injected wastes have properties that can be toxic to living organisms.

(b) ***Reactivity***

Reactivity describes a waste's tendency to interact chemically with other substances. Many wastes are reactive, but it is the degree of reactivity that defines a waste as hazardous. Hazardous reactive wastes are those that are normally unstable and readily undergo violent change without detonating, react violently with water, form potentially explosive mixtures with water, generate toxic gases, or fumes when combined with water, contain sulfide or cyanide and are exposed to extreme pH conditions, or are explosive. Because deep-well-injected waste streams are usually dilute (typically less than 1% waste in water) hazardous reactivity is not a significant consideration in deep-well injection, although individual compounds may exhibit this property at higher concentrations than those that exist in the waste stream. Nonhazardous reactivity is, however, an important property in deep-well injection, because when a reactive waste is injected, precipitation reactions that can lead to well plugging may occur.

(c) ***Corrosivity***

Corrosive wastes are defined as those wastes with a pH $\leq 2 \geq 12.5$ (i.e., the waste is very acidic or very basic). Beyond its importance in defining a waste as hazardous, the corrosivity of wastes is also a property of concern to deep-well injection systems and operations. Corrosive wastes may damage the injection system, typically by electrochemical or microbiological means. Corrosion of injection well pumps, tubing, and other equipment can lead to hazardous waste leaking into strata not intended for injection. For information on various types of electrochemical corrosion relevant to the injection-well system, the reader is referred to (57). Other recommended sources include (58–62), which discuss saturation and stability indexes for predicting the potential for corrosion or scaling (accumulation of carbonate and sulfate precipitates) in injection wells. The Stiff and Davis Index is recommended by Warner and Lehr (57) as most applicable to deep-well injection of hazardous wastes, because it is intended for use with highly saline ground waters. Additionally, Ostroff (63) provides examples of how to use the index, Watkins (64) describes procedures that test for corrosion, and Davis (65) thoroughly discusses microbiological corrosion of metals.

(d) ***Ignitability***

As noted, deep-well-injected wastes are relatively dilute. Therefore, ignitability is not a significant consideration in deep-well injection, although in a concentrated form, individual compounds may exhibit this property. Ignitability has no further implications for the fate of deep-well-injected waste.

8.2. Sources, Amounts and Composition of Injected Wastes

The sources, amounts, and composition of injected hazardous wastes are a matter of record, since the Resource Conservation and Recovery Act (RCRA) requires hazardous waste to be manifested (i.e., a record noting the generator of waste, its composition or characteristics, and its volume must follow the waste load from its source to its ultimate disposal site). The sources and amounts of injected hazardous waste can be determined, therefore, based on these records. Table 14.4 shows the estimated volume of deep-well-injected wastes by industrial category in 1983 (56). More than 11 billion gallons of total hazardous waste were injected in 1983. Organic chemicals (51%) and petroleum-refining and petrochemical products (25%) accounted for three-quarters of the volume of injected wastes. The remaining 24% was divided among six other industrial categories: miscellaneous chemical products, agricultural chemical products, inorganic chemical products, commercial disposal, metals and minerals, and aerospace and related industry.

Although the general composition of each shipment of wastes to an injection well may be known, a number of factors make it difficult to characterize fully the overall composition of industrial wastewaters at any one well. These factors include (1) variations in flow, in concentrations, and in the nature of organic constituents over time; (2) biological activity that may transform constituents over time; and (3) physical inhomogeneity (soluble and insoluble compounds) (66). Further, the exact composition of the shipment may not be known because of chemical complexity (66). An example of the complexity of organic wastes is illustrated in

Table 14.4
Estimated volumes of deep-well injected wastes by industrial category, 1983

Industrial category	Volume (MG/year)	Percent of total
Organic chemical	5868	50.9
Petroleum refining and petrochemical products	2888	25.0
Miscellaneous chemical products	687	6.0
Agricultural chemical products	525	4.6
Inorganic chemical products	254	2.2
Commercial disposal	475	4.1
Metals and minerals	672	5.8
Aerospace and related industry	169	1.5
Total	11, 539	100.0

Roy et al. (67), which presents an analysis of an alkaline pesticide-manufacturing waste. This waste contained more than 50 organic compounds, two fifths of which could not be precisely identified.

Although no systematic data base exists on the exact composition of deep-well-injected wastes in a survey of 209 operating waste-injection wells, Reeder et al. (68) found that 53% injected one or more chemicals identified in that study as hazardous. The US EPA gathered data for 108 wells (55% of total active wells) that were operated in 1983. Table 14.5 summarizes the total quantity of undiluted waste in six major categories, provides a breakdown of average concentrations of constituents for which data were available, and indicates the number of wells involved. A little more than half the undiluted waste volume was composed of nonhazardous inorganics (52%). Acids were the most important constituent by volume (20%), followed by organics (17%). Heavy metals and other hazardous inorganics made up less than 1% of the total volume in the 108 wells. About a third of the wells injected acidic wastes and about two thirds injected organic wastes. Although the percentage of heavy metals by volume was low, almost one fifth of the wells injected wastes containing heavy metals.

An injected wastestream typically is composed of the waste material and a large volume of water. Because the data in Table 14.5 include only 55% of the injection wells that were active in 1983, it is not possible to estimate precisely the percentage of waste to the total volume of injected fluid shown in Table 14.4. However, if the same total proportions apply to all wells, wastes made up of 3.6% of the total volume of injected fluid (36,000 mg/L). This percentage agrees well with an independent estimate for a typical injection ratio of 96% water and 4% waste (69).

Table 14.5 also shows that the average concentration of all the acidic wastes exceeded 40,000 mg/L. Concentrations of metals ranged from 1.4 mg/L (chromium) to 5,500 mg/L (unspecified metals, probably containing multiple species). Five of the 18 organic constituents exceeded 10,000 mg/L (total organic carbon, organic acids, formaldehyde, chlorinated organics, and formic acid); four exceeded 1,000 mg/L (oil, isopropyl alcohol, urea nitrogen, and organic peroxides).

Table 14.5
Waste characteristics of 108 hazardous waste wells active in 1983 in the United States

Waste Type/ Components	Gallons[a]	Average Concentration (mg/L)	No. of Wells
Acids	44,140,900 (20.3)[b]		35 (32.4)[b]
Hydrochloric acid		78,573	15
Sulfuric acid		43,000	6
Nitric acid		75,000	2
Formic acid		75,000	2
Acid, unspecified		44,900	12
Heavy metals	1,517,600 (0.7)		19 (17.6)
Chromium		1.4	11
Nickel		600	5
Metals, unspecified		5,500	2
Metal hydroxides		1,000	1
Hazardous inorganics	89,600 (<0.1)		4 (3.7)
Selenium		0.3	2
Cyanide		391	2
Organics	39,674,500 (17.4)		71 (65.7)
Total organic carbon (TOC)		11,413	24
Phenol		805	22
Oil		3,062	6
Organic acids		10,000	3
Organic cyanide		400	3
Isopropyl alcohol		1,775	3
Formaldehyde		15,000	2
Acetophenone		650	2
Urea “N”		1,250	2
Chlorinated organics		35,000	2
Formic acid		75,000	2
Organic peroxides		4,950	2
Pentachlorophenol		7.6	2
Acetone		650	2
Nitrile		700	1
Methacrylonitrile		22	1
Ethylene chloride		264	1
Carbon tetrachloride		970	1
Nonhazardous Inorganics	118,679,700 (52.0)	–	50 (46.3)
Other	22,964,600 (9.9)	–	33 (30.5)
Total	228,021,800[c]		108

[a] Gallons of nonaqueous wastes before dilution and injection.
[b] Number in parentheses is the percentage of total.
[c] Excludes overlaps between organics and acids.

8.3. Geographic Distribution of Wells

The use of wells for disposal of industrial wastes dates back to the 1960s, when it was implemented primarily in response to more stringent water pollution control regulations (70). The number of industrial-waste injection wells more than doubled between 1967 and 1986.

The state distribution shows some interesting patterns. Class I injection wells are concentrated in two states, Texas (112 wells) and Louisiana (70 wells), which have a total of 69% of all wells in 1986. The growth from 1984 to 1986 has been concentrated in Texas, with a 38% increase, from 81 to 112 wells. The only other states to show a significant increase from 1984 to 1986 are Indiana (13 proposed wells) and California (7 proposed wells). Nine states have had industrial-waste injection wells in the past but did not have any permitted Class I wells in 1986 (Alabama, Colorado, Iowa, Mississippi, Nevada, North Carolina, Pennsylvania, Tennessee, and Wyoming). One state (Washington) had a Class I well in 1986, but no record of industrial wastewater injection before that year.

The 1989 US EPA data show that the heavy concentration of hazardous waste injection wells is in three geologic basins: Gulf Coast, Illinois Basin, and the Michigan Basin.

8.4. Design and Construction of Wells

The following description of the design and construction of deep-injection wells is adapted from Donaldson (71), Donaldson et al. (72), and US EPA (73).

8.4.1. Surface Equipment Used in Waste Disposal

Figure 14.7 shows the surface equipment used in a typical subsurface waste-disposal system. Detailed discussion of surface treatment methods can be found in Warner and Lehr (57). The individual elements are:

(a) A *sump tank* or an open 30,000- to 50,000-gal steel tank is commonly used to collect and mix waste streams. An oil layer or, in a closed tank, an inert gas blanket is often used to prevent air contact with the waste. Alternatively, large, shallow, open ponds may provide sufficient detention time to permit sedimentation of particulate matter. Such ponds often are equipped with cascade, spray, or forced-draft aerators to oxidize iron and manganese salts to insoluble forms that precipitate in the aeration ponds.

(b) An *oil separator* is used when the waste contains oil because oil tends to plug the disposal formation. The waste is passed through a settling tank equipped with internal baffles to separate the oil from the waste.

(c) A *clarifier* removes such particulate matter as polymeric flocs, dirt, oil, and grease. It is often a tank or a pond in which detention time is long enough to allow suspended particles to settle gradually. The process also may be accelerated by adding a flocculating agent such as aluminum sulfate, ferric sulfate, or sodium aluminate. Tank clarifiers are often equipped with a mechanical stirrer, sludge rake, and surface skimmer that continuously remove sludge and oil.

(d) A *filter* is used in some cases when coagulation and sedimentation do not completely separate solids from the liquid waste in areas where sand and sandstone formations are susceptible to plugging. Filters with a series of metal screens coated with diatomaceous earth or cartridge filters typically are used. Where limestone formations with high solution porosity are used for injection, filtration is usually not required.

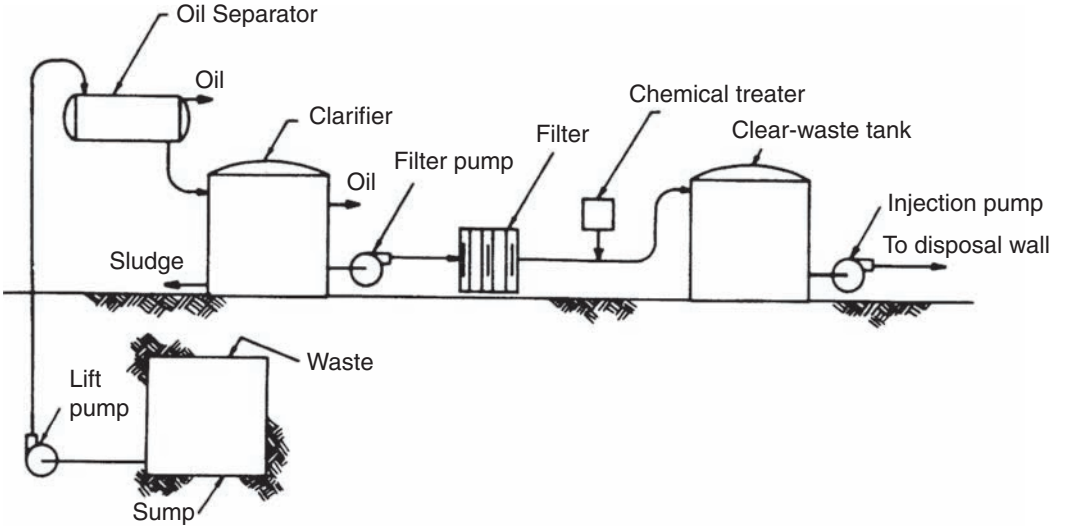


Fig. 14.7. Above-Ground components of a subsurface waste disposal system. (*Source*: US EPA).

(e) A *chemical treater* is used to inject a bactericide if microorganisms could cause fouling of injection equipment and plugging of the injection reservoir.
(f) An unlined steel *clear-waste tank* typically is used to hold clarified waste before injection. The tank is equipped with a float switch designed to start and stop the injection pump at predetermined levels.
(g) An *Injection pump* is used to force the waste into the injection zone, although in very porous formations, such as cavernous limestone, the hydrostatic pressure of the waste column in the well is sufficient. The type of pump is determined primarily by well-head pressures required, the volume of liquid to be injected, and the corrosiveness of the waste. Single-stage centrifugal pumps are used in systems that require wellhead pressures up to about 150 psi, and multiplex piston pumps are used to achieve higher injection pressures.

8.4.2. Injection-Well Construction

Most injection wells are drilled using the rotary method, although depending on the availability of equipment and other site-specific factors, reverse- rotary or cable-tool drilling may be used. The construction of an injection well incorporates several important elements: (a) bottom-hole and injection-interval completion, (b) casing and tubing, (c) packing and cementing, (d) corrosion control, and (e) mechanical-integrity testing. A detailed discussion of the technical aspects of industrial-waste injection-well construction can be found in Warner and Lehr (57). US EPA (73) also presents a survey of well construction methods and materials used for 229 hazardous waste injection wells.

1. Two types of *Injection well completions* are used with hazardous waste injection wells:
   (a) *Open hole* completion typically is used in competent formations such as limestone, dolomite, and consolidated sandstone that will stand unsupported in a borehole. In 1985, 27% of Class I wells were of this type, with most located in the Illinois Basin.

(b) *Gravel pack* and *perforated* completions are used where unconsolidated sands in the injection zone must be supported. In gravel-pack completions the cavity in the injection zone is filled with gravel or, more typically, a screen or liner is placed in the injection-zone cavity before the cavity is filled with gravel. In perforated completions, the casing and cement extend into the injection zone and are then perforated in the most permeable sections. In 1985, 53% of Class I wells were perforated and 17% were screened (73).

2. *Casing* and *tubing* are used to prevent the hole from caving in and to prevent aquifer contamination by confining wastes within the well until they reach the injection zone. Lengths of casing of the same diameter are connected together to form casing strings. Usually two- or three-casing strings are used. The outer casing seals the near-surface portion of the well (preferably to below the point where aquifers containing less than 10,000 mg/L total dissolved solids, potential underground sources of drinking water, are located). The inner casing extends to the injection zone. Tubing is placed inside the inner casing to serve as the conduit for injected wastes, and the space between the tubing and casing is usually filled with kerosene or diesel oil after packing and cementing are completed.
3. *Packers* are used at or near the end of the injection tubing to plug the space, called the annulus, between the injection tubing and the inner casing. Cement is applied to the space between the outer walls of the casing and the borehole or other casing. Portland cement is used most commonly for this purpose, although when acidic wastes are injected, special acid-resistant cements are sometimes used in the portion of the well that passes through the confining layers.
4. *Corrosion control* can be handled several ways: (a) by using corrosion-resistant material in constructing the well, (b) by treating the waste stream through neutralization or other measures, and (c) by cathodic protection.
5. *Mechanical Integrity testing* is required by EPA regulations to ensure that an injection well has been constructed or is operating without (a) significant leakage from the casing, tubing, or packer or (b) upward movement of fluid through vertical channels adjacent to the well bore. Types of mechanical integrity tests include the following:
    1. Pressure test
    2. Monitor annual pressure
    3. Temperature log
    4. Noise log
    5. Radioactive tracer log
    6. Cement bond log
    7. Caliper log
    8. Casing condition log

A detailed discussion of mechanical integrity can be found in US EPA (74).

8.5. Disposal of Radioactive Wastes

Radioactive contaminants created by nuclear fission and other mean differ from the usual industrial plant waste in their ability to emit radiation. There is no known method for neutralizing radioactivity, but radioactive isotopes decay and thus lose their activity with time.

The Halliburton Company Inc. has developed (British Patent L 054740) an improved method for the disposal of a radioactive waste by mixing the waste with cement to form slurry then injecting it through a well. A horizontal fracture is developed hydraulically and the slurry is injected into the opening and caused to harden in place. This method, which has been used successfully since about 1967 at Oak Ridge, TN, is available for disposal of

low and intermediate level radioactive wastes. In using this method, a conventional injection well is drilled; generally to a depth less than 1,000 ft but until an impermeable formation is transversed. The nonpermeable formation is perforated by, for example, gun perforations. A fracturing fluid, which may be the waste-cement slurry, is pumped into the well under sufficient pressure (greater than I psi/ft of depth) to exceed the formation breakdown (fracture) pressure. The formation will fracture or part generally in a horizontal direction at this depth. The waste-cement slurry is injected into the fracture and the fracture is sealed by allowing the cement to harden.

The same procedure then may be repeated at other depths within the well. To cause the cement to harden, the slurry may need an absorption type clay such us attapulgite. At other times, an agent such as calcium lignosulfonate that reacts chemically with the cement and retards its setting time may need to be added.

Another little used but technically sound method for the disposal of radioactive wastes is the injection through well radioactive or toxic materials into strata of low permeability at depth of 1,000 to 20,000 ft, or at such a depth that the wastes are removed from the biosphere. This method involves displacement of formation connate water with liquid waste. Rocks of low permeability that are impregnated with waste become permanent storage receptors if the differential pressures across the receptor strata are maintained at a minimum. They will be retained almost indefinitely as a film held by molecular adhesion on the wall of the interstices (75–77).

Other factors also play a role in the retention of fluids in formations of low permeability. For example, the greater the amount of total interstitial surface in a rock or unconsolidated material, the greater is its specific retention. As would be expected, it is found that, as the effective diameter of the material's grain decreases, the specific retention generally increases because the total exposed surface area increases with decreasing grain size.

Capillarity is important in the retention of fluids in any granular material such as sedimentary rocks. The openings between the granules are interconnected in all directions, with the result that capillary forces act out in all directions within such materials. The moisture film around particles is held so tightly that it strongly resists any forces tending to displace it. The degree of its resistance to movement is expressed by its capillary potential, which is a measure of the force required to move this moisture from the soil.

Ion exchange is an important geochemical process. Many ions in hazardous waste products maybe removed from wastewaters by means of this process. Clay minerals exhibit a marked capacity for the exchange of cations; in fact, all clay formations possess some ion exchange capacity. Clay minerals exhibiting good ion exchange are: kaolinite, halloysite, montmorillonite, illite, vermiculite, chlorite, sepiolite, attapulgite, and polygorskite. Of these, the montmorillonites are noted for the highest cation-exchange capacity, and the kaolins are noted for the most rapid rate of exchange.

The cohesive property of fluids plays an important part in their retention or movement in porous rocks Cohesion is the ability of a fluid, or other substance to stick together and resist separation.

The adhesive properties of a rock also play an important part in the movement or retention of a fluid in a porous rock. Adhesion is a measure of a fluid's ability to stick to the surfaces of other materials, such as to rock material in a formation.

The salty water found in deeply buried sedimentary formation generally is ancient seawater called connate water, Understanding the openings or pore spaces of the rock materials that build up on the ocean floor during geologic time and contain the connate water is the key to understanding the value of these natural spaces as permanent receptors for storage of hazardous and radioactive wastes.

9. PROTECTION OF USABLE AQUIFERS

Any aquifer that contains water that is both economically practical and technologically feasible to use for domestic, agricultural, industrial, or other purposes should be protected. Such aquifers are vulnerable to contamination by either the injected fluids or fluids being displaced by the injected fluids migrating into the aquifer to be protected. A variety of measures described below can be used to assure that injection well systems will not contaminate protected aquifers (78).

To protect groundwater, the well must be constructed so as to prevent contamination by (a) keeping injected fluids within the injection well casing and within the intended injection zone, and (b) keeping formation fluids displaced by the injected fluids from migrating into a protected aquifer. There are six major pathways (79) by which fluids may move and contaminate aquifers. The following discussion describes each pathway and summarizes a way to prevent migration through that pathway.

9.1. Pathway 1: Migration of Fluids through a Faulty Injection Well Casing

The casing of a well can serve a variety of purposes. It supports the well bore to prevent collapse of the geologic formations into the hole and consequent loss of the well, it serves as the conductor of injected fluids from the land surface to the intended injection zone and supports other components of the well. If a well casing is defective, injected fluids may escape through the defect and enter the protected aquifer (Fig. 14.8). Such migration can contaminate an underground source of drinking water.

To detect migration of fluids through leaks in the casing periodic tests of the casing's integrity should be made. Several types of casing inspection tests are commercially available. For example, the downhole TV camera (80) can "see" what is wrong on the inside of the casing. The downhole casing inspection log called Vertilog (81) is a system for making a quantitative measurement of corrosive damage, indicating if the metal loss is internal or external and if it is isolated or circumferential. Holes in the casing can be identified as well as casing separations.

Use of separate tubing for injection affords protection to the casing and decreases the possibility of leakage. It isolates the casing of the well from injected fluids. By preventing this contact between casing and injected fluids, the possibility of migration of contaminants through leaks in the casing is greatly diminished. For the same reason, the use of tubing and packer also lessens the chances of corrosion of casing. By monitoring the annular space

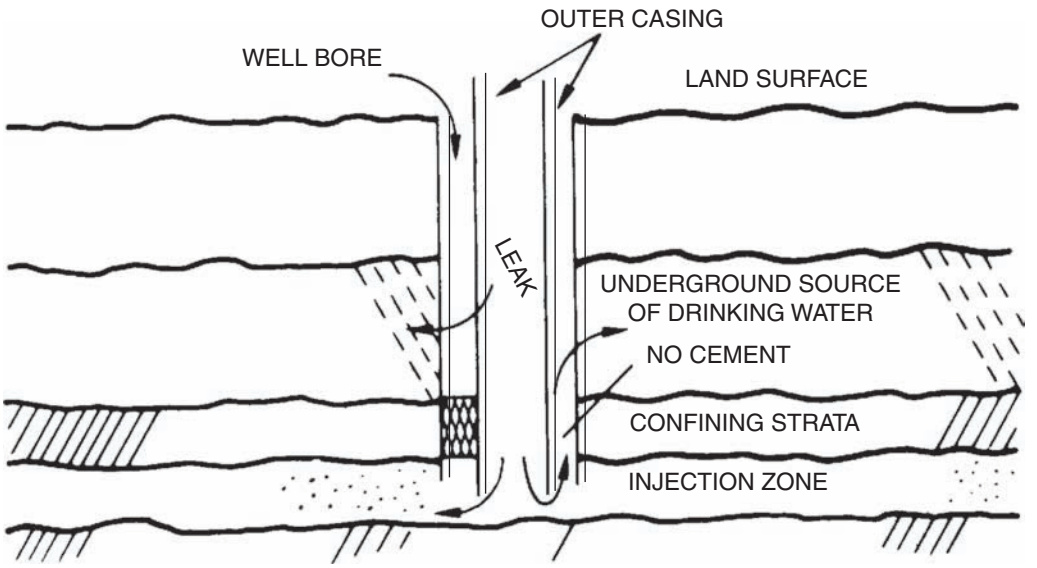


Fig. 14.8. Faulty well construction. (*Source*: US EPA).

between the casing and tubing, leaks in the tubing can be detected and repaired before the casing becomes faulty. Tubing and packer offers two further advantages. It isolates the annulus (between the tubing and casing) from the injection zone, facilitating detection of any leaks in the tubing. It allows for visual inspection for deterioration of the tubing during routine maintenance. Finally, wells that inject corrosive fluids should be constructed of corrosion-resistant materials. This material is intended to prolong the operating life and continued viability of the well casing.

9.2. Pathway 2: Migration of Fluids Upward Through the Annulus between the Casing and the Well Bore

A second pathway by which contaminants can enter protected aquifers is by migrating upward through the annulus, between the drilled hole and the casing. Under usual injection conditions, injected fluids, upon leaving the well, enter a stratum in the injection zone that to some degree resists the entry of the fluids. This resistance results from friction and is inversely proportional to the size of the small openings in the stratum. Because fluids tend to take the path of least resistance, unless properly contained, they may travel upward through this annulus. If sufficient injection pressure exists, the fluids could migrate upward through the drilled hole into the overlying protected aquifer.

Leaks through holes in the well casing or upward fluid movement between the well's outer casing and the well bore, are illustrated in Fig. 14.8.

Casing should be cemented to isolate the aquifers to be protected from all underlying saline aquifers and from the injection zone. Generally two 100-ft thick cement plugs are installed. One is located immediately below the lowermost aquifer to be protected. The other is located

immediately above the injection zone. The absence of leaks and fluid movement in the well bore should be confirmed periodically using geophysical logging techniques.

9.3. Pathway 3: Migration of Fluids from an Injection Zone through the Confining Strata

The third way by which fluids can enter a protected aquifer is through leaks in the confining strata. Upon entry into an injection zone, fluids injected under pressure will normally travel away from the well and laterally through the receiving formation. In most cases, this occurrence is expected and gives rise to no concern, but, if the confining stratum that separates the injection zone from an overlying or underlying protected aquifer leaks significantly because it is either fractured or permeable, the injected fluids can migrate out of the receiving formation and into the protected aquifer.

For obvious reasons, there is no general well construction standard that can address this problem of migration of fluids through the confining Strata.

Several steps should be taken to assure that fluids do not migrate through or around the confining strata (Fig. 14.9).

1. Select a deep formation as a receptor. The deeper the receptor stratum selected for injection, the greater the degree of protection.
2. Place at least 200 ft of cement about the injection zone. The thicker the cement plug placed above the injection zone the larger the pathway of fluid movement before flow can enter the well bore above the plug.
3. Study the confining and receptor strata. Care should be taken during drilling of the test hole to determine the permeability, thickness, and other information for the confining and injection strata, and the changes in fluid chemistry that can be expected as the fluids migrate through the confining bed.
4. Determine fluid movement rates at various pressures. The leakage rate versus injection pressure should be evaluated before operation, and the injection pressure limited to avoid fluid movement through the confining strata into protected aquifers.

Frequently, when leakage out of the receptor stratum occurs, the adjacent aquifers (those leaked into) are permeable enough so that only limited vertical migration occurs. The equation for pressure buildup in the injection shell as a result of injecting into a zone with a leaky confining bed on one side is (82),

$$P_{\mathrm{r}} = P_i + P_{\mathrm{DL}}[141.2Q\mu B]/\mathrm{k}H \tag{5}$$

$$P_{\mathrm{DL}} = \text{a function of } (t_D, r/B) \tag{6}$$

$$t_{\mathrm{D}} = 6.33 \times 10^{-3} kT/\Phi\mu Cr^2 \tag{7}$$

$$B = \sqrt{kHh_{\mathrm{c}}/k_{\mathrm{c}}} \tag{8}$$

The equation for determination of pressures in the injection well where both confining beds leak is the same as given above except that:

$$B = \sqrt{kHh_{\mathrm{c}}h_{\mathrm{c}}{}'/k_{\mathrm{c}}{}'h_{\mathrm{c}} + k_{\mathrm{c}}h_{\mathrm{c}}{}'} \tag{9}$$

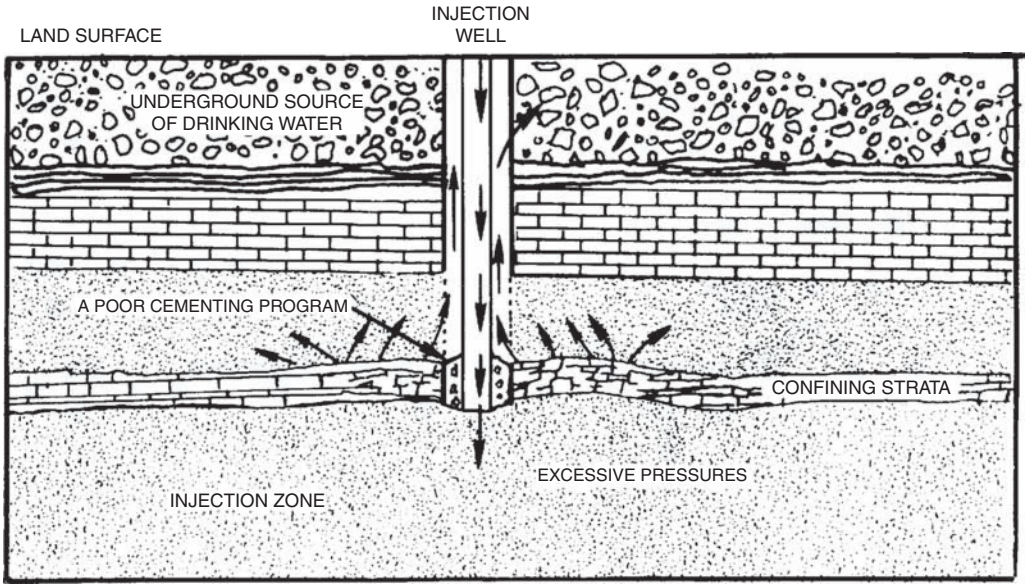


Fig. 14.9. Leakage through confining strata. (*Source*: US EPA).

where:

P_r = reservoir pressure at radius r (ft of water or psi)
P_i = initial formation pressure (ft of water or psi)
P_{DL} = dimensionless pressure for semiconfined reservoirs
t_D = dimensionless time
B = leakage factor
t_D = dimensionless time
r = radial distance from well bore (ft)
k = average permeability (millidarcy)
k_c = vertical permeability of confining layer (millidarcy)
k_c' = vertical permeability of second confining layer (millidarcy)
H = reservoir thickness (ft)
h_c = thickness of confining layer (ft)
h_c' = thickness of second confining layer (ft)
C = compressibility $(\text{psi})^{-1}$
Φ = porosity expressed as a decimal
μ = viscosity

For examples of solutions for Eqs. (5–9), the reader is referred to the US EPA publication 600/2-79-170 (83).

9.4. Pathway 4: Vertical Migration of Fluids through Improperly Abandoned or Improperly Completed Wells

Fluids from the pressurized area in the injection zone may be forced upward through nearby wells (Fig. 14.10) that penetrate the injection horizon within a zone around the injection well called the zone of endangering influence. The zone of endangering influence of a well may be defined as that radial distance from the well bore within which pressure increases because of injection are sufficient to cause a potential for upward migration into freshwater zones. As shown in Fig. 14.11, the zone of endangering influence includes all that area surrounding an injection well wherein the upward pressure in the injection zone exceeds the downward pressure of freshwater when measured using the base of freshwater as the datum. In areas where before injection the upward pressures in the injection formation already exceed the downward freshwater pressure, the zone of endangering influence is infinite or very large. In such areas an alternate fixed radius of 0.25 mile (84) was approved by US EPA for the review of nearby wells.

Wells located within the area of pressure increase from an injection well should be examined to assure that they are properly completed and plugged. Corrective action should be taken where necessary to prevent fluids from migrating along these pathways into protected aquifers.

The key aspects influencing zone pressures during injection are:

(a) The existing fluid pressures in the disposal aquifer
(b) The pressure increases induced to effect waste fluid emplacement

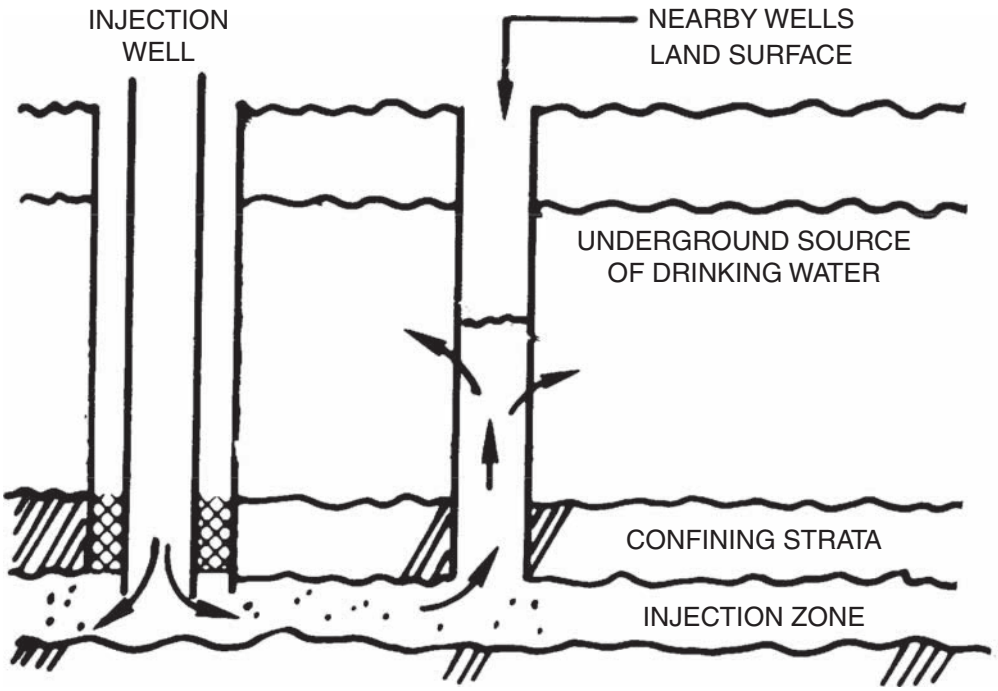


Fig. 14.10. Leakage through nearby wells. (*Source*: US EPA).

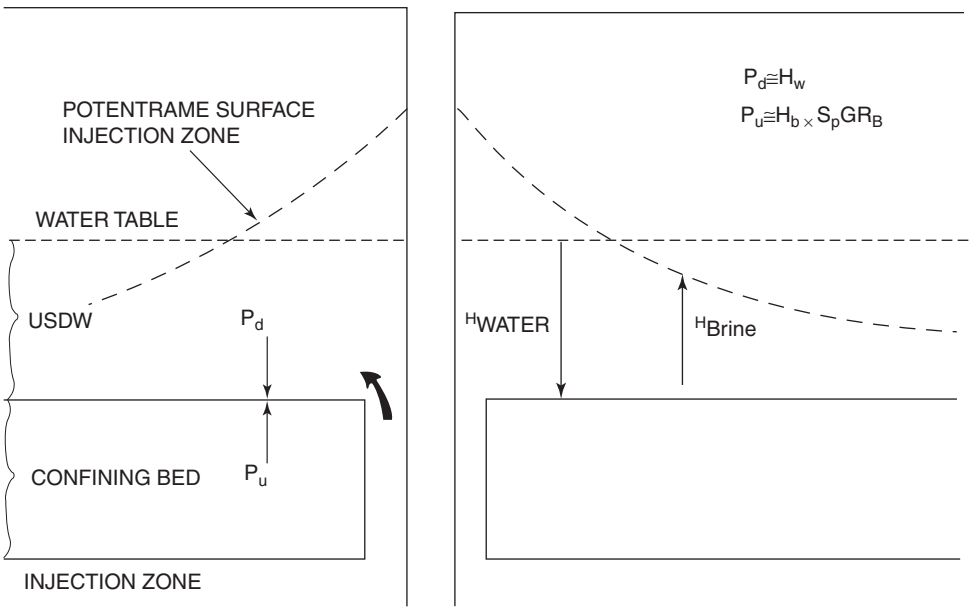


Fig. 14.11. Zone of endangering influence. Where $P_u > P_d$, then potential for endangerment exists. (*Source*: US EPA).

Several methods common to the practice of engineering hydrology and reservoir analysis are applicable to the solution of the zone of endangering influence (Fig. 14.11). These methods analyze the pressure differential that exists at the base of freshwater between increased reservoir pressures because of injection, and the pressure exerted downward by the freshwater column.

Consider an example. Increased formation pressures cause a column of water to rise in an open hole to a level 100 ft above the base of freshwater. Thus at the base of freshwater, the formation fluid is exerting an upward pressure equal to 100 ft of hydrostatic head (of brine). However, if the freshwater aquifer contains, say, 200 ft of water, then a downward pressure of about 200 ft of head (of freshwater) is exerted at the base. Under these conditions if leakage were to occur the freshwater would "leak" downward into the brine. Only when upward pressure (P_u) is greater than the downward pressure (P_d) can there be the potential for upward movement.

The zone of endangering influence around an injection well encompasses all the area within which pressure increases because of injection are sufficient to create an upward differential pressure, measured at the base of freshwater.

This type analysis assumes "worst case" (i.e., open hole) conditions. An analysis considering friction losses through small channels, drilling mud column displacement or seepage through beds would yield a much smaller "zone."

The pressure change in an injection zone at distance (r) caused by injection volume (Q_t) may be described by the equation (85) for predicting pressure increases given below:

$$s = 162.6(Qu/kH)\log(kt/70.4\Phi uCr^2) \tag{10}$$

where:

s = change in pressure (ft of water or psi)
Q = flow or injection rate (barrels/day)
$u = 1.87r^2S/Tt$ (centipoises = cp)
k = average permeability (millidarcy)
H = reservoir thickness (ft)
t = time (days)
Φ = porosity expressed as a decimal
C = compressibility $(\text{psi})^{-1}$
r = radial distance from well bore (ft)

This pressure increase is additive to the existing formation pressure before injection began (P_2).

Step 1:

Solve Eq. (7) for any two values of r, and convert s to feet of hydrostatic head by dividing (psi) by the gradient per foot of the formation fluid (Table 14.6). Add this value to P_2 (hydrostatic head of the injection zone) and plot the two values at their respective r on semilog paper. A straight line connecting the two points establishes the pressure surface of the disposal zone as it exists in space, measured in feet of head of formation brine above the top of the injection zone.

Similarly, some pressure (P_1) exists in the basal freshwater aquifer, corresponding to the weight of a column of water in a well fully penetrating that aquifer.

Step 2:

Locate the stratigraphic position of the base of the lowermost freshwater aquifer on the diagram. Convert (feet of head of freshwater) to (feet of head as formation water) measured from the base of the freshwater aquifer. Draw a horizontal line to denote the pressure surface of the freshwater as it exists in space.

Step 3:

The intersection (if any) of lines P_1 and $(s + P_2)$ denotes the radius (r) of the "zone of endangering influence." That is, to the left of the intersection the pressure in the disposal zone is sufficient to overcome the hydrostatic pressure at the base of freshwater, and an upward potential is realized.

This method of solving the "zone of endangering influence" of an injection well represents an extremely conservative viewpoint.

Table 14.6
Density of Brines

Specific gravity of water at 60°F	Approximate total solids in parts per million, mg/L	Weight in psi of pressure per ft[a]
1.000	none	0.433
1.010	13,500	.437
1.020	27,500	.441
1.030	41,400	.445
1.040	55,400	.450
1.050	69,400	.454
1.060	83,700	.459
1.070	98,400	.463
1.080	113,200	.467
1.090	128,300	.471
1.100	143,500	.476
1.110	159,500	.480
1.120	175,800	.485
1.130	192,400	.489
1.140	210,000	.493

[a]Averages approximately 0.0043 psi/ft for each increase of 0.01 indensity.

9.5. Pathway 5: Lateral Migration of Fluids from Within an Injection Zone into a Protected Portion of Those Strata

In most cases, the injection zone of a particular well will be physically segregated from underground sources of drinking water by impermeable material. In some instances, however, wells inject into an unprotected portion of an aquifer that is hydraulically interconnected with a protected aquifer (Fig. 14.12). In this event, there may be no impermeable layer or other barrier to prevent contact between contaminated fluids and underground drinking water.

Injection into unprotected portions of aquifers that contain drinking water in other areas must be done with great care. This type of injection can work if the predominant flow of the aquifer is such that injected fluids will tend to move away from, rather than toward, the protected part of the aquifer.

It is sometimes helpful to define the actual position of the waste front and its movement with lime.

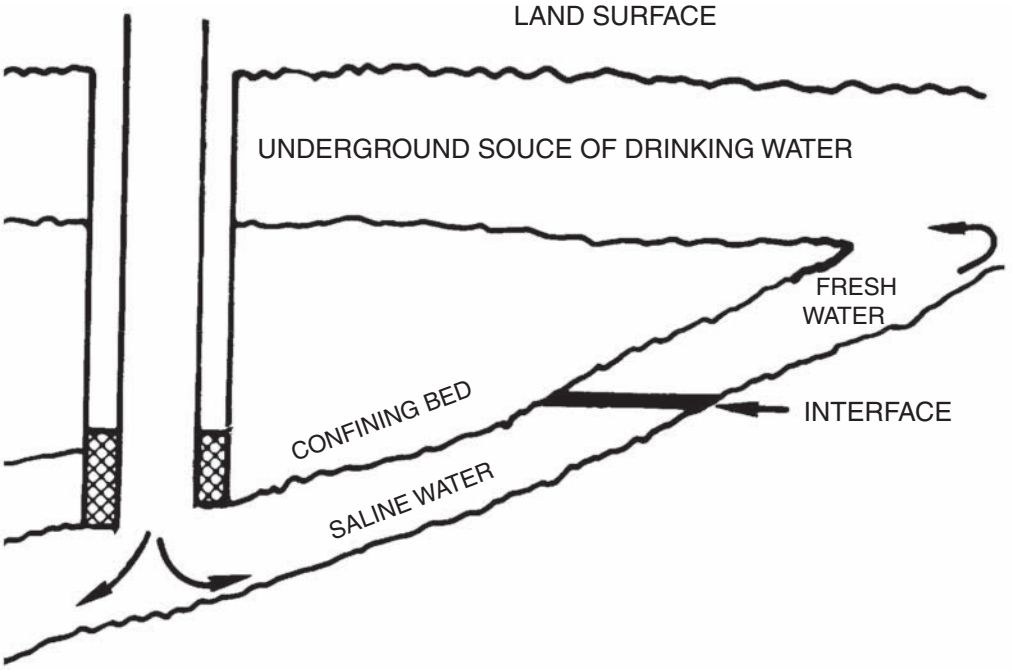


Fig. 14.12. Hydraulically interconnected aquifers. (*Source*: US EPA).

The minimum distance the waste will have traveled during injection may be described (86) by

$$r = (V/\pi H\Phi)^{1/2} \tag{11}$$

where

r = radial distance from well bore (ft)
V = injected volume, (ft^3)
H = reservoir thickness (ft)
Φ = porosity expressed as a decimal

A typical solution is given in of Section 11.5.

In most practical situations, the minimum radial distance of travel will be exceeded because of dispersion, density segregation, and channeling through higher permeability zones.

An estimate of the influence of dispersion can be made with the Eq. (9)

$$\gamma = r + 2.3\sqrt{Dr} \tag{12}$$

where:

γ = radial distance from well bore with dispersion, ft

r = radial distance from well bore (ft)
D = dispersion coefficient

This equation is obtained by solving Eq. (6.5) of Bear (87) for the radial distance at which the injection front has a chemical concentration of 0.2%. A dispersion coefficient of three represents a reasonable value for a sandstone aquifer (see Section 11.6).

It may be impossible to accurately predict the chemical character of the plume of waste 100 years after it has slowly moved a few hundred feet in contact with subsurface minerals. However, it is important to list some of the chemical and biological reactions that will certainly occur to degrade the waste. The rates at which these reactions will occur are only speculative, however. These reactions include (75, 76, 88–93):

1. Dilution and dispersion
2. Biological degradation of organic compounds
3. Biochemical degradation of some species, such as nitrate, sulfate, and so on
4. Adsorption and ion exchange reactions with clay particles
5. pH neutralization
6. Precipitation and immobilization of some constituents

It is much more important to attempt to predict the direction and ultimate location of the waste front. As was pointed out in an earlier section, the disposal reservoir is confined above and below by a number of relatively impermeable, regionally persistent clay formations. Confined by these rocks, the disposal reservoir dips and thickens toward the Gulf of Mexico, In other words, the further the waste moves coastward, the deeper and more separated from freshwater it becomes. Near the coastline, the formation exists at depths exceeding 10,000 ft. At this point, the angle of dip steepens radically, and the formation dips beneath the Gulf of Mexico to depths exceeding 20,000 ft.

Therefore, over a period of millions of years, confined above and below by clay harriers and separated from freshwater by thousands of feet of rocks, a gradually decomposed waste will move at a very slow rate and remain essentially isolated in the deep subsurface.

9.6. Pathway 6: Direct Injection of Fluids into or Above an Underground Source of Drinking Water

The last pathway of contamination of groundwater is also the most hazardous. Direct injection of contaminated fluids into or above underground sources of drinking water presents the most immediate risk to public health. Such direct injection causes an instantaneous degradation of groundwater (Fig. 14.13). The injected fluids do not benefit from natural treatment processes such as filtration and ion exchange.

Many shallow injection wells, pits, septic tanks, and other similar disposal systems arc used to dispose contaminants above drinking water aquifers that need to be protected (94, 95). The injected fluids then percolate downward into drinking water aquifers, as illustrated in Fig. 14.10. US EPA (84) decided that wells injecting hazardous wastes directly into drinking water aquifers are be banned. Drilling of new wells was prohibited after 1982.

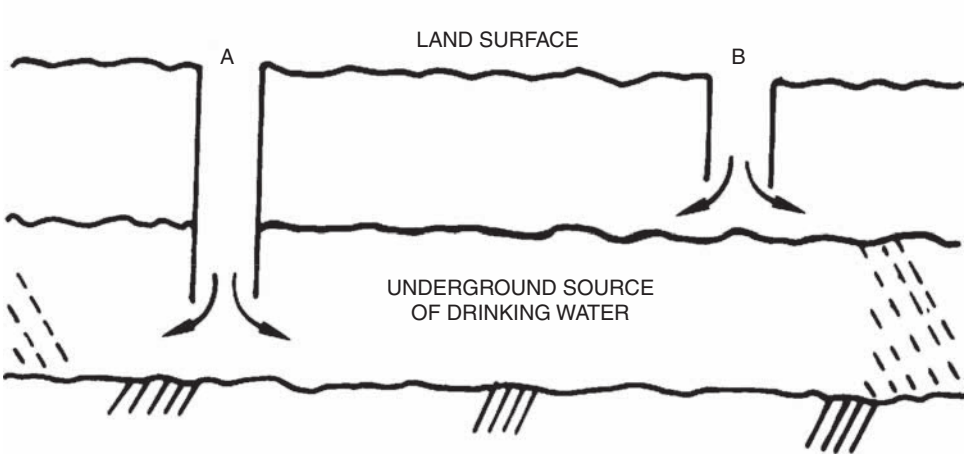


Fig. 14.13. Direct injection. (*Source*: US EPA).

Casing should be installed through all aquifers to be protected and cemented to isolate them from exposure to fluids being injected and from all underlying saline aquifers penetrated by the well bore.

10. CASE STUDIES OF DEEP WELL INJECTION

Field studies are an important complement to geochemical modeling and to laboratory studies. Two ways to investigate interactions between injected wastes and reservoir material are (a) direct observation of the injection zone and overlying aquifers using monitoring wells and (b) backflushing the injected waste (56). In both instances, samples of the fluids in the zone are collected at intervals to characterize the nature of geochemical reactions and to track changes over time.

1. *Monitoring Wells*. Monitoring wells drilled into the injection zone at selected distances and directions from the injection well allow direct observation of formation water characteristics and the interactions that occur when the waste front reaches the monitoring well. When placed near the injection well in the aquifer above the confining layer, monitoring wells can detect the upward migration of wastes caused by casing or confining-layer failure. Foster and Goolsby (96) describe detailed methods for constructing monitoring wells. Monitoring wells have several advantages: time-series sampling of the formation over extended periods is easy and the passage of the waste front can be observed precisely. Disadvantages are cost and the potential for upward migration of wastes if monitoring well casings fail. A monitoring well at the Monsanto plant had to be plugged when unneutralized waste reached it because of fears that the casing would corrode (*see* Section 10.1.1). The two Florida case studies (Sections 10.1 and 10.2) and the North Carolina case study (Section 10.3) illustrate the usefulness of monitoring wells.
2. *Backflushing of Injected Wastes*. Backflushing of injected wastes can also be a good way to observe *waste*/reservoir geochemical interactions. Injected wastes are allowed to backflow (if formation pressure is above the elevation of the wellhead) or are pumped to the surface. Backflowed

wastes are sampled periodically (and reinjected when the test is completed); the last sample taken will have had the longest residence time in the injection zone. Keely (97) and Keely and Wotf (98) describe this technique for characterizing contamination of near-surface aquifers and suggest using logarithmic time intervals for chemical sampling. The two Florida studies (Sections 10.1 and 10.2) all present results from backflushing experiments. The advantages of backflushing are reduced cost compared with that of monitoring wells and reduced sampling time (sampling takes place only during the test period). Disadvantages include less precise time- and distance-of-movement determinations and the need to interrupt injection and to have a large enough area for backflushed fluid storage before reinjection.

10.1. Case Study 1: Pensacola, FL (Monsanto)

10.1.1. Injection-Facility Overview

Monsanto operates one of the world's largest nylon plants on the Escambia River about 13 miles north of Pensacola, Florida. The construction, operations, and effects of the injection-well system at this site have been extensively documented by the US Geological Survey in cooperation with the Florida Bureau of Geology. Pressure and geochemical effects are reported by Goolsby (99). Additional microbiological data are reported by Elkan and Horvath (100). Major chemical processes observed at the site include neutralization, dissolution, biological denitrification, and methanogenesis. The geochemical fate of organic contaminants in the injected wastes, however, has not been reported.

The waste is an aqueous solution of organic monobasic and dibasic acids, nitric acid, sodium and ammonium salts, adiponitrile, hexamethylenediamine, alcohols, ketones, and esters (99). The waste also contain cobalt, chromium, and copper, each in the range of 1 to 5 mg/L. Waste streams with different characteristics, produced at various locations in the nylon plant, are collected in a large holding tank; this composite waste is acidic. The specific characteristics of the waste varied somewhat as a result of process changes. Until mid-1968, wastes were partially neutralized by pretreatment. After that, unneutralized wastes were injected. No reason was reported for suspending treatment. Goolsby (99) reports pH measurements ranging from a high of 5.6 in 1967 to a low of 2.4 in 1971, and Eh ranging from +300 mV in 1967 to +700 mV in 1971. The chemical oxygen demand in 1971 was 20,000 mg/L.

Monsanto began injecting wastes into the lower limestone of the Floridan aquifer in 1963. In mid-1964, a second well was drilled into the formation about 1,000 ft southwest of the first. A shallow monitoring well was placed in the aquifer above the confining layer about 100 ft from the first injection well, and a deep monitoring well was placed in the injection zone about 1,300 ft south of both injection wells. The deep monitoring well (henceforth referred to as the near- deep monitoring well) was plugged with cement in 1969. In late 1969 and early 1970, two additional deep monitoring wells were placed in the injection formation, 1.5 miles south-southeast (downgradient) and 1.9 miles north-northwest (upgradient) of the site. From 1963 to 1977, about 13.3 billion gallons of waste were injected. During the same period, injection pressures ranged from 125 to 235 psi.

Ten months after injection of neutralized wastes began; chemical analyses indicated that dilute wastes had migrated 1300 ft to the nearest deep monitoring well. Injection of unneutral-

ized wastes began in April 1968. Approximately 8 months later, unneutralized wastes reached the near-deep monitoring well, indicating that the neutralization capacity of the injection zone between the injection wells and the monitoring well had been exceeded. At this point, the monitoring well was plugged with cement from bottom to top because operators were concerned that the acidic wastes could corrode the steel casing and migrate upward (99). The rapid movement of the waste through the limestone indicated that most of it migrated through a more permeable section, which was about 65 ft thick. By mid-1973, 10 years after injection began, a very dilute waste front arrived at the south monitoring well, 1.5 miles away. As of early 1977, there was no evidence that wastes had reached the upgradient monitoring well. The shallow monitoring well remained unaffected during the same period.

Increases in permeability caused by limestone dissolution approximately doubled the injection index (the amount of waste that can be injected at a specified pressure). As of 1974, the effects of the pressure created by the injection were calculated to extend more than 40 miles radially from the injection site. An updip movement of the freshwater/saltwater interface in the injection-zone aquifer, which lies less than 20 miles from the injection wells, was also observed.

10.1.2. Injection/Confining-Zone Lithology and Chemistry

The lower limestone of the Floridan aquifer is used as the injection zone (at 1,400 to 1,700 ft), and the Bucatunna clay member of the Byram formation (about 220 ft thick) serves as the confining layer. Figure 14.14 shows the local stratigraphy and the monitoring well installations. The formation water in the injection zone is a highly saline (11,900 to 13,700 mg/L total dissolved solids, TDS) sodium-chloride solution. The Eh of samples collected from two monitoring wells located in the injection formation ranged from +23 to −32 mV, indicating reducing conditions in the injection zone that would favor anaerobic biodegradation.

The injection zone contains about 7,900 mg/L chloride, but less than 20 miles northeast of the injection site, chloride concentrations are less than 250 mg/L. Under natural conditions, water in the injection zone moves slowly south-southwestward toward the Gulf of Mexico, where it is assumed to discharge about 100 miles offshore. The preinjection hydraulic gradient was about 1.3 ft/mile.

10.1.3. Chemical Processes Observed

As a result of dissolution of the limestone by the partly neutralized acid wastes, calcium concentrations more than doubled in the near-deep monitoring well 10 months after injection started in 1963 (99). In early 1966, however, they dropped to background levels (about 200 mg/L), possibly in response to biochemical decomposition of the waste. In September 1968, after about 300 MG of the acidic, unneutralized waste had been injected, the calcium concentration began to increase again. An abrupt increase in calcium to 2,700 mg/L, accompanied by a decrease in pH to 4.8 in January 1969 led to the decision to plug the near-deep monitoring well.

In an attempt to find out how fast the waste was reacting with limestone, a 3-hour backflushing experiment, in which waste was allowed to flow back out of the injection well,

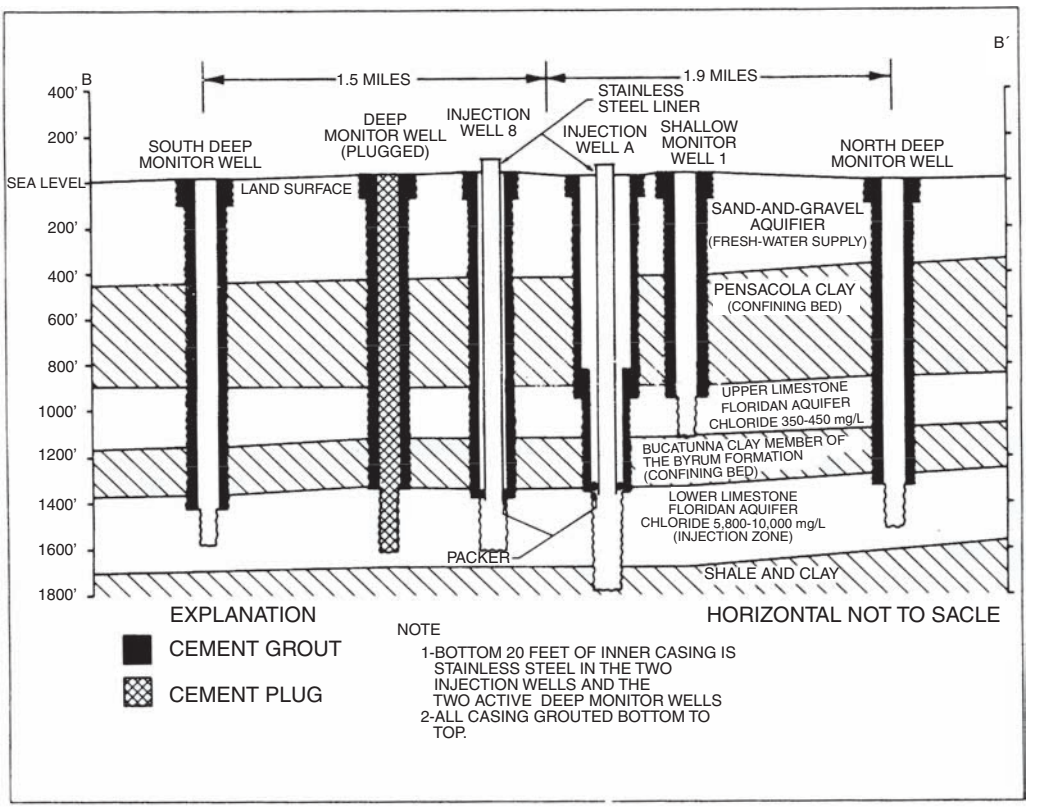


Fig. 14.14. Pensacola injection facility hydrogeologic cross-section. (*Source*: US EPA).

yielded some unexpected results. The increase in pH of the neutralized waste could not be fully accounted for by the solution of limestone as determined from the calcium content of the backflushed liquid: the additional neutralization apparently resulted from reactions between nitric acid and alcohols and ketones in the original waste induced by increased pressure in the injection zone compared to surface conditions.

The lack of nitrates (which were present at levels of 545 to 1,140 mg/L in the waste) in the near-deep monitoring well, combined with the presence of nitrogen gas, indicated that degradation by denitrifying bacteria had taken place (99). Backflushing shortly before injecting unneutralized wastes confirmed denitrification. Nitrate concentrations decreased rapidly as the backflushed waste was replaced by formation water. Similar backflushing experiments conducted after unneutralized wastes were injected, however, provided no evidence of denitrification, indicating that microbial activity was suppressed in the portion of the zone containing unneutralized wastes.

Elkan and Horvath (100) performed a microbiological analysis of samples taken from the north and south deep monitoring wells in December 1974; about 6 months after the dilute waste front had reached the south well. Both denitrifying and methanogenic bacteria

were observed. The lower numbers and species diversity of organisms observed in the south monitoring well compared with those in the north well indicated suppression of microbial activity by the dilute wastes.

Chemical analyses of the north and south monitoring wells were published for the period March 1970 to March 1977. Between September 1973 and March 1977 bicarbonate concentrations increased from 282 to 636 mg/L and dissolved organic carbon increased from 9 to 47 mg/L. These increases were accompanied by an increase in the dissolved-gas concentration and a distinctive odor like that of the injected wastes. The pH, however, remained unchanged. During the same period, dissolved methane increased from 24 to 70 mg/L, indicating increased activity by methanogenic bacteria. The observation of denitrification in the near-deep monitoring well and methanogenesis in the more distant south monitoring well fit the redox-zone biodegradation model.

Significant observations made at this site are: (a) organic contaminants (as measured by dissolved organic carbon) continue to move through the aquifer even when acidity has been neutralized, and (b) even neutralized wastes can suppress microbial populations.

10.2. Case Study 2: Belle Glade, FL

10.2.1. Injection-Facility Overview

The Belle Glade site, located southeast of Lake Okeechobee in south-central Florida, illustrates some of the problems that can develop with acidic-waste injection when carbonate rock is the confining layer. Contributing factors to the contamination of the aquifer above the confining zone were the dissolution of the carbonate rock and the difference in density between the injected wastes and the formation fluids. The injected waste was less dense than the ground water because of its lower salinity and higher temperature (101).

The injected fluids include the effluent from a sugar mill and the waste from the production of furfural, an aldehyde processed from the residues of processed sugar cane. The waste is hot (about 75° to 93 °C); acidic (pH 2.6 to 4.5); and has high concentrations of organics, nitrogen, and phosphorus. The waste is not classified as hazardous under 40 CFR 261, and the well is currently regulated by the State of Florida as a nonhazardous injection well. The organic carbon concentration exceeds 5,000 mg/L.

The well was originally cased to a depth of 1,495 ft, and the zone was left as an open hole to a depth of 1,939 ft. The depth of the zone has been increased twice. Seasonal injection (fall, winter, and spring) began in late 1966; the system was inactive during late summer. Injection rates ranged from 400 to 800 gpm, and wellhead injection pressures ranged from 30 to 60 psi. By 1973 injection had become more or less continuous. From 1966 to 1973, more than 1.1 billion gallons of waste were injected (101).

At the time injection began, a shallow monitoring well was placed 75 ft south of the injection well in the upper part of the Floridan aquifer above the confining layer. A downgradient, deep monitoring well was placed in the injection zone 1000 ft southeast of the injection well. Another shallow well, located 2 miles southeast of the injection site at the University of Florida's Everglades Experiment Station, has also been monitored for near-surface effects.

Acetate ions from the injected waste were detected in the deep monitoring well 1,000 ft southeast of the injection well in early 1967; a matter of months after injection began. In 1971, about 27 months after injection began; evidence of waste migration was detected at a shallow monitoring well in the upper part of the Floridan aquifer. Dissolution of the carbonate confining layer by the acidic waste was the main reason for the upward migration. However, the lower density of the injected wastes compared with that of the formation waters (0.98 vs. 1,003 g/mL) served to accelerate the rate of upward migration (101). In an attempt to prevent further upward migration, the injection well was deepened to 2,242 ft, and the inner casing was extended and cemented to 1938 ft. When waste injection was resumed, evidence of upward migration to the shallow aquifer was observed only 15 months later. By late 1973, 7 years after injection began; the waste front was estimated to have migrated 0.6 to 1 mile from the injection well.

The injection well was deepened a third time, to a depth of 3,000 ft. A new, thicker confining zone of dense carbonate rock separates the current injection zone from the previous zone (see Fig. 14.15-the current injection zone is not shown). As of early 1989, the wastes were still contained in the deepest injection zone.

10.2.2. Injection/Confining-Zone Lithology and Chemistry

The wastes are injected into the lower part of the carbonate Fioridan aquifer, which is extremely permeable and cavernous (see Fig. 14.15). The natural direction of groundwater flow is to the southeast. The confining layer is 150 ft of dense carbonate rocks. The chloride concentration in the upper part of the injection zone is 1,650 mg/L, increasing to 15,800 mg/L near the bottom of the formation (101). The sources used for this case study did not provide any data on the current injection zone. The native fluid was basically a sodium-chloride solution but also included significant quantities of sulfate (1,500 mg/L), magnesium (625 mg/L), and calcium (477 mg/L).

10.2.3. Chemical Processes Observed

Neutralization of the injected acids by the limestone formation led to concentrations of calcium, magnesium, and silica in the waste solution that were higher than those in the unneutralized wastes. Anaerobic decomposition of the organic matter in the injected waste apparently occurred through the action of both sulfate-reducing and methanogenic bacteria. Sulfate-reducing bacteria were observed in the injected wastes that were allowed to backflow to the surface. Sulfate levels in the native ground water declined by 45%, and the concentration of hydrogen sulfide increased by 1,600%. Methane fermentation (reduction of CO_2 to CH_4) was also inferred from the presence of both gases in the backflow fluid, but the presence of methanogenic bacteria was not confirmed. Increased hydrogen sulfide concentrations produced by the bacteria during biodegradation and the subsequent decrease in sulfate/chloride ratio in the observation wells were taken as indicators of upward and lateral migration. Migration into the shallow monitoring well was also indicated by a decline in pH from around 7.8 to 6.5, caused by mixing with the acidic wastes.

Chemical analyses of the backflowed injected waste that had been in the aquifer for about 2.5 months (for which some dilution had occurred) indicated that chemical oxygen demand

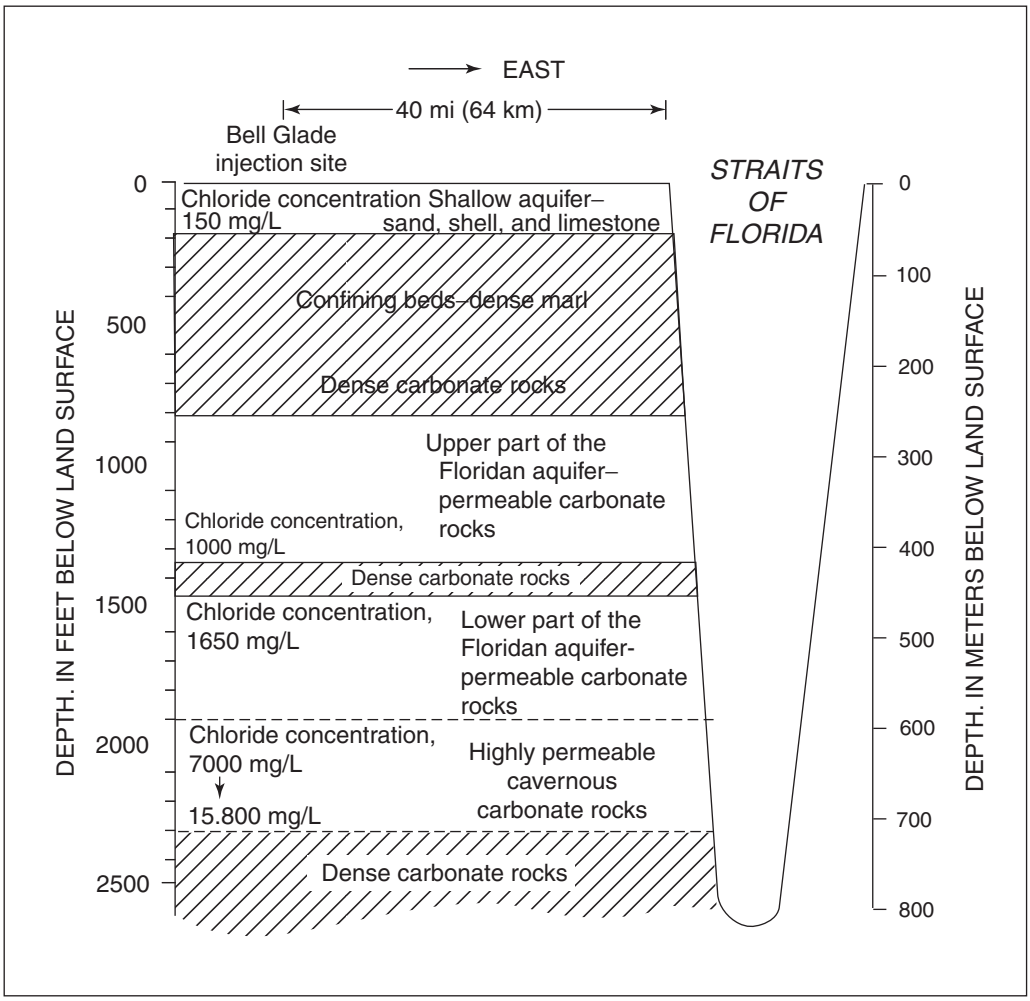


Fig. 14.15. Generalized hydrogeologic section between Belle Glade and the straights of Florida. (*Source*: US EPA).

(COD) was about half that of the original waste. Samples that had been in residence for about 5 months had a COD approximately one quarter that of the original waste (12,200 mg/L in the original waste compared with 4,166 mg/L in the samples). The percent reduction in COD resulting from bacterial action rather than dilution was not estimated.

10.3. Case Study 3: Wilmington, NC

10.3.1. Injection-Facility Overview

The Hercules Chemical, Inc. (now Hercufina, Inc.), facility, 4 miles north of Wilmington, North Carolina, attempted deep-well injection of its hazardous wastes from May 1968 to

December 1972, but had to discontinue injection because of waste-reservoir incompatibility and unfavorable hydrogeologic conditions. The US Geological Survey conducted extensive geochemical studies of this site until the well was abandoned (102, 103). Biodegradation processes were also studied (100, 104). More geochemical- fate processes affecting injected organic wastes have been documented at this site than at any other.

Hercules Chemical produced an acidic organic waste derived from the manufacture of dimethyl terphthalate, which is used in the production of synthetic fiber. The average dissolved organic carbon concentration was about 7,100 mg/L and included acetic acid, formic acid, *p*-toluic acid, formaldehyde, methanol, terphthalic acid, and benzoic acid. The pH ranged from 3.5 to 4.0. The waste also contained traces (less than 0.5 mg/L) of 11 other organic compounds, including dimethyl phthalate, a listed hazardous waste.

From May 1968 to December 1972, the waste was injected at a rate of about 300,000 gpd. The first injection well was completed to a depth of 850 to 1,025 ft (i.e., cased from the surface to 850 ft with screens placed in the most permeable sections of the injection zone to a depth of 1,025 ft). One shallow observation well was placed 50 ft east of the injection site at a depth of 690 ft. Four deep monitoring wells were also placed in the injection zone, one at 50 ft and three at 150 ft from the injection well.

The injection well became plugged after a few months of operation because of the reactive nature of the wastes and the low permeability of the injection zone. The actual plugging process was caused both by reprecipitation of the initially dissolved minerals and by plugging of pores by such gaseous products as carbon dioxide and methane. When the first well failed, a second injection well was drilled into the same injection zone about 5,000 ft north of the first, and injection began in May 1971. Nine additional monitoring wells (three shallow and six deep) were placed at distances ranging from 1,500 to 3,000 ft from the second injection well. Injection was discontinued in 1972 after the operators determined that the problems of low permeability and waste- reservoir incompatibility could not be overcome. Monitoring of the waste movement and subsurface environment continued into the mid-1970s in the three monitoring wells located 1,500 to 2,000 ft from the injection wells.

Within 4 months, the waste front had passed the deep observation wells located within 150 ft of the injection well. About 9 months after injection began; leakage into the aquifer above the confining layer was observed. This leakage was apparently caused by the increased pressures created by formation plugging and by the dissolution of the confining beds and the cement grout surrounding the well casing of several of the deep monitoring wells, caused by organic acids.

Eight months after injection began in the second injection well, wastes had leaked upward into the adjacent shallow monitoring well. The leak apparently was caused by the dissolution of the cement grout around the casing. In June 1972, 13 months after injection began in the second well, the waste front reached the deep monitoring well located 1,500 ft northwest of the injection well, and in August 1972 waste was detected in Well about 1,000 ft north of injection well. Waste injection ended in December 1972. As of 1977, the wastes were treated in a surface facility (100).

10.3.2. Injection/Confining-Zone Lithology and Chemistry

The injection zone consisted of multiple Upper Cretaceous strata of sand, silty sand, clay, and some thin beds of limestone (see Fig. 14.16). The clay confining layer was about 100 ft thick. The total-dissolved- solids concentration in the injection-zone formation water was 20,800 mg/L, with sodium chloride the most abundant constituent.

10.3.3. Chemical Processes Observed

A number of chemical processes were observed at the site (102, 103):

1. The waste organic acids dissolved carbonate minerals, alumino-silicate minerals, and iron/manganese-oxide coatings on the primary minerals in the injection zone.
2. The waste organic acids dissolved and formed complexes with iron and manganese oxides. These dissolved complexes reprecipitated when the pH increased to 5.5 or 6.0 because of neutralization of the waste by the aquifer carbonates and oxides.
3. The aquifer mineral constituents adsorbed most waste organic compounds except formaldehyde. Adsorption of all organic acids except phthalic acid increased with a decrease in waste pH.
4. Phthalic acid was complexed with dissolved iron. The concentration of this complex decreased as the pH increased because the complex coprecipitated with the iron oxide.
5. Biochemical waste transformation occurred at low waste concentrations, resulting in the production of methane. Additional microbial degradation of the waste resulted in the reduction of sulfates to sulfides and ferric ions to ferrous ions.

When the dilute waste front reached the monitoring well, in June 1972, microbial populations rapidly increased in this well, with methanogenesis being the major degradative process (104). Elkan and Horvath (100) found greater numbers and species diversity of microorganisms in one of the wells, which contained dilute wastes, than in the well, which was uncontaminated. In laboratory experiments, however, DiTommaso and Elkan (104) found that bacterial growth was inhibited as the concentration of waste increased and could not decompose the waste at the rate it was being injected.

This case study illustrates the importance of dissolution/precipitation reactions in determining waste- reservoir compatibility. Adsorption was observed to immobilize most of the organic constituents in the waste except for formaldehyde. As with the Monsanto case study, biodegradation was an important process when wastes were diluted by formation waters, but the process became inhibited when undiluted waste reached a given location in the injection zone.

11. PRACTICAL EXAMPLES

11.1. Example 1

The rate of leakage through a confining bed can be determined from

$$Q = PIA \tag{1}$$

where:

Q = the rate of leakage, ft^3/day

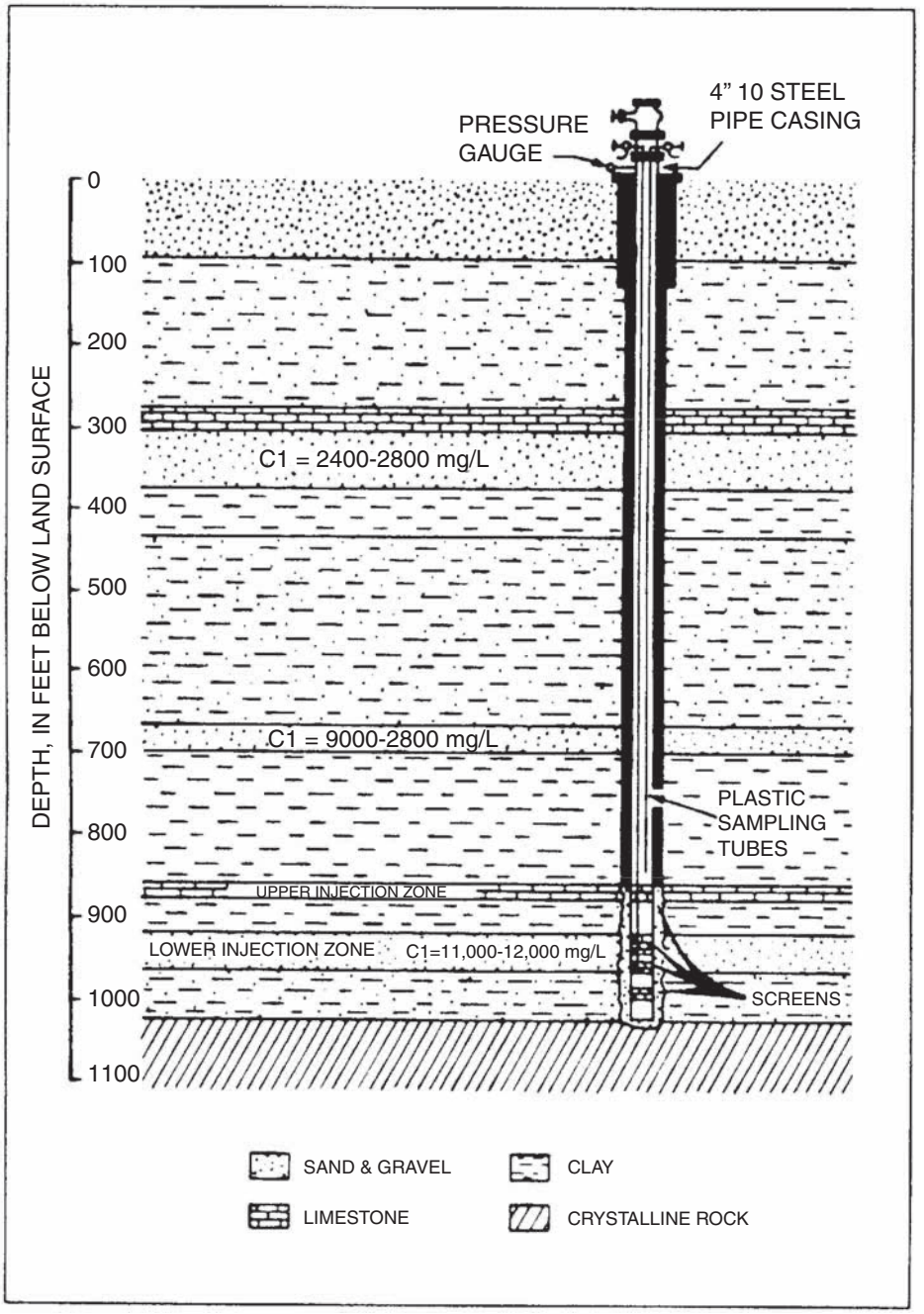


Fig. 14.16. Construction features and lithologic log of well 14, Wilmington, NC. *Source*: US EPA).

P = Permeability, $ft^3/day/ft^2$
I = Hydraulic gradient, ft/ft
A = Area, ft^2

If one assumed that the permeability (P) of a typical clay confining bed averaged 0.002 gal/day/ft^2 (93×10^{-12} L/s/cm^2) or (0.00027 $ft^3/day/ft^2$), that the difference in hydraulic pressure across the confining bed is 50 ft (15.24 in.) and that the confining bed is 50 ft thick (Hydraulic gradient I of 1ft/ft), then the leakage (Q) calculated through the confining bed for a circular area (A) within 1000 ft (305 m) radius of the injection well would be:

$P = 0.00027\ ft^3/day/ft^2$
$I = 50\ ft/50\ ft = 1$
$A = \pi(1000)^2$
$Q = (0.00027)(1)(3.142 \times (1000)^2 = 848\ ft^3/day$

11.2. Example 2

The average velocity of fluid movement through the confining bed in Example 1 is determined from:

$$v = PI/\Phi \tag{2}$$

where

P = permeability = 0.00027 $ft^3/day/ft^2$
I = hydraulic gradient = 1 ft/ft
Assume ϕ = porosity = 0.25
Then $v = 0.00027 \times 1/0.25 = 0.001$ ft/day

At a velocity of 0.001 ft/day, the waste front would move through a 50 ft thick confining bed in 50,000 days (137 year).

11.3. Example 3

Assume that the fluid pressure in an observation well located 500 ft (152 m) from a well injecting wastewater at a rate of 500 gpm (32 L/s) was recorded for about a week. Values of r^2/t are computed, and then plotted against changes in pressure in the observation well for different values t. These are superimposed upon a second graph sheet that values of $W(u)$ and u from Table 14.2 had previously been plotted on at the same scale. The coincident value of the match point at $u = 0.01$ and $W(u) = 1.0$, are $r^2/t = 2.8 \times 10^5$ and s = 5.5 ft (Fig. 14.5). Solving for T and S gives:

$$T = \frac{114.60QW(u)}{s} \tag{3}$$

$$S = uT/1.87(r^2/t) \tag{4}$$

$T = 114.60(500)(1.00)/(5.5) = 10,420\,\text{gal/day/ft}$
$S = (0.01)(10,420)/1.87\ (2.8 \times 10^5) = 0.0002$

Having determined the T and S coefficient, long-range pressure changes with time for various distances can be forecast by rearranging the above formulas thus:

$$u = 1.87r^2S/Tt \tag{5}$$

$$s = 114.60QW(u)/T \tag{6}$$

For example, the pressure increase at a point of 5,000 ft (1524 m) from an injection site after injecting 800 gpm for 5 years (1,825 days) in the above described receptor would he:

$$u = (1.87)(5000)(5000)(0.0002)/(10,420 \times 1825) = 0.0005$$

From Table 14.2, $W(u) = 7.02$

$$s = (114.60)(800)(7.02)/10,420 = 62\text{ft}$$

Note: This method is not valid for $u > 0.02$.

11.4. Example 4

Consider the following example of how to determine pressure increases from injection which are sufficient to cause an upward gradient in the injection zone, when considered in an open hole at the base of freshwater.

1. At a distance, $r = 10$ ft
2. At a distance $r = 100$ ft

Given that,

$Q = 15,000$ barrels/day
$u = 1$centipoise (cp)
$k = 1127$ millidarcy (md)
$H = 300$ ft
$t = 7,300$ days (20 years)
$\phi = 0.346$
$C = 6.5 \times 10^{-6}$/psi

The pressure change in an injection zone at distance (r) is given by s:

$$s = 162.6(Qu/kH)\log(kt/70.4\Phi uCr^2) \tag{7}$$

The solution for pressures change at $r = 10$ ft and $r = 100$ ft in the receptor zone are given below:

1. $s_{(r=10)} = 162.6\ (15,000)\ (1)/1127)\ (300)\ \log\ (1127)\ (7300)/70.4\ (.346)(1)\ (6.5 \times 10^{-6})(10)^2 = 63$ psi
2. $s_{(r=100)} = 162.6\ (15,000)\ (1)/1127)\ (300)\ \log\ (1127)\ (7300)/70.4\ (0.346)\ (1)\ (6.5 \times 10^{-6})\ (100)^2 = 49$ psi

11.5. Example 5

The minimum distance (r) the waste will have traveled during injection may be described by (86):

$$r = (V/\pi H \Phi)^{1/2} \tag{8}$$

Therefore for,

$Q = 15,000$ barrels/day, @ 42 gal/barrel $= 630,000$ gpd $= 84,\ 000\,\text{ft}^3/\text{day}$
$H = 300\,\text{ft}$
$\phi = 0.346$

and for $t = 5$ years will yield a volume $V = 84,000 \times 5 \times 365 = 1.53 \times 10^8\,\text{ft}^3$

$$\text{r} = [(1.53 \times 10^8)/(3.14 \times 300 \times .0.346)]^{1/2} = 685\,\text{ft}$$

and for $t = 20$ years will yield a volume $V = 84,000 \times 20 \times 365 = 6.12 \times 10^8\,\text{ft}^3$

$$r = [(6.12 \times 10^8)/(3.14 \times 300 \times 0.346)]^{1/2} = 1370\,\text{ft}$$

11.6. Example 6

In most practical situations, the minimum radial distance of travel will be exceeded because of dispersion, density segregation, and channeling through higher permeability zones.

An estimate of the influence of dispersion ($D = 3$) can be made with the equation:

$$\gamma = r + 2.3\sqrt{Dr} \tag{9}$$

for $r = 685\,\text{ft}$,
$\gamma = 685 + 2.3\sqrt{3 \times 685} = 790\,\text{ft}$
for $r = 1370\,\text{ft}$,
$\gamma = 1370 + 2.3\sqrt{3 \times 1370} = 1517\,\text{ft}$

NOMENCLATURE

A = area (ft^2)
B = leakage factor
C = compressibility $(\text{psi})^{-1}$
D = dispersion coefficient
H = reservoir thickness (ft)
h_c = thickness of confining layer (ft)
h_c' = thickness of second confining layer (ft)
I = hydraulic gradient (ft/ft)

k = average permeability (millidarcy)
k_c = vertical permeability of confining layer (millidarcy)
k_c' = vertical permeability of second confining layer (millidarcy)
L = leakage factor for semiconfined aquifer = $\sqrt{khh_c/k_c}$
P = coefficient of permeability (gal/day/ft^2)
P_{DL} = dimensionless pressure for semiconfined reservoirs
P_i = initial formation pressure (ft of water or psi)
P_1 = hydrostatic pressure in the base of freshwater (ft of water or psi)
P_2 = hydrostatic pressure in the injection zone (ft of water or psi)
P_r = reservoir pressure at radius r (ft of water or psi)
P_u = upward pressure
P_d = downward pressure
Q = flow or injection rate (ft^3/day, gpm or barrels/day)
r = radial distance from well bore (ft)
s = change in pressure (ft of water or psi)
S = coefficient of storage
t = time (day)
t_D = dimensionless time
T = transmissibility (gal/day/ft)
$u = 1.87r^2S/Tt$ (centipoises = cp)
$V = Q\,t$ = cumulative volume of waste injected (ft^3)
v = fluid velocity (ft/day)
$W(u)$ = well function of u given in Table 14.1
β = formation volume factor = $\frac{\text{Volume of liquid at reservoir temperature and pressure}}{\text{Volume of liquid at standard temperature and pressure}}$
Φ = porosity expressed as a decimal
γ = radial distance from well bore with dispersion (ft)
μ = viscosity
$\pi = 3.14$

REFERENCES

1. US EPA, *Technical Program Review: Underground Injection Control Regulations*, US Environmental Protection Agency, Office of Water, EPA 816-R-02-025, December (2002).
2. US EPA, *What is the UIC Program? Summary of the History of the UIC Program*, US Environmental Protection Agency URL: http://www.epa.gov/safewater/uic/whatis.html, August (2004).
3. US EPA, *EPA Deep Wells (Class I)*, US Environmental Protection Agency URL: http://www.epa.gov/safewater/uic/classi.html, November (2002).
4. GWPC, *Injection Well Bibliography by Ground Water Protection Council*, 3rd ed., Ground Water Protection Council, Oklahoma City, Oklahoma, August (1995).
5. D. R. Rima, E. B. Chase, and B. M. Myers, *Subsurface Waste Disposal by Means of Wells-A Selected Annotated Bibliography*, US Geological Survey, Water-Supply Paper 2020 (1971).

6. US EPA, *Compilation of Industrial and Municipal Injection Wells in the United States*, US Environmental Protection Agency, October (1974).
7. D. L. Warner, Survey of Industrial Waste Injection Wells, vols. I, II, and III, US Department of Commerce, AD756 642, June (1972).
8. E. C. Donaldson, Injection wells and operations today, in *Underground Waste Management and Environmental Implications*, The American Association of Petroleum Geologists, Tulsa, Oklahoma (1972).
9. D. L. Warner and O. H. Orcutt, Industrial wastewater-injection wells in United States-Status of use and regulations, in *Underground Waste Management and Artificial Recharge*, American Association of Petroleum Geologists, Vol. 2, Tulsa, OK (1973).
10. A. Sellinger and S. H. Aberhach, Artificial recharge of coastal-plain aquifer in Israel. In: *Underground Waste Management and Artificial Recharge*, American Association of Petroleum Geologists, Vol. 2, Tulsa, OK (1973).
11. A. Amramy, *Civil Engineering*, 38, 5, 58 *May* (1968).
12. E. C. Donaldson and A. F. Bayayerd, *Reuse and Subsurface Injection of Municipal Sewage Effluent-Two Case Histories*, US Bureau of Mines, Information Circular 8522 (1971).
13. US EPA, *Ocean Outfalls and Other Methods of Treated Wastewater Disposal in Southeast Florida*, US Environmental Protection Agency, Final Environmental Impact Statement, March 19 (1973).
14. The Congress of the United States of America, Public Law 92–500 known as the *Federal Water Pollution Control Act Amendments of 1972*, 33 USC 1151 (1972).
15. R. O. Bernon, The Beneficial Uses of Zones of High Transmissivities in the Florida Subsurface for Water Storage and Waste Disposal, Florida Department of Natural Resources, Information Circular 70 (1970).
16. US EPA, *What is Underground Injection?*, US Environmental Protection Agency, URL: http://www.epa.gov/safewater/uic/history.html, March (2004).
17. J. J. Hickey and J. Vecchioli, *Subsurface Injection of Liquid Waste with Emphasis on Injection Practices in Florida*., US Geological Survey, Water Supply Paper 2281 (1984).
18. US EPA, Subsurface emplacement of fluids, US Environmental Protection Agency, Administrators Decision Statement #5, *Fed. Reg*., 39, 69, 12922–3, April 9 (1974).
19. Geraghty and Miller, Inc., *The Ground Water Environment* (1973).
20. B. Knape, *Underground Injection Operations in Texas: A Classification and Assessment of Underground Injection Activities*, Report 291, Texas Department of Water Resources, Austin, TX, December (1984).
21. US EPA, *Class I Underground Injection Control Program: Study of the Risks Associated with Class I Underground Injection Wells*, EPA 816-R-01-007, US Environmental Protection Agency, Office of Water, March (2001).
22. CMA, *Class I Underground Injection Wells: Responsible Management of Chemical Wastes*, Chemical Manufacturers Association, Pamphlet (1994).
23. US EPA, *Mid-Course Evaluation of the Class II Underground Injection Control Program*, Final Report of the Workgroup, Underground Injection Control Branch, US Environmental Protection Agency, Office of Drinking Water, August (1989).
24. Engineering Enterprises, *Analysis of Casing and Cementing Compliance Policy for Class II Wells*, EPA Contract No. 68-03-3416, US EPA, US Environmental Protection Agency, Office of Drinking Water, October (1987).

25. GAO, *Safeguards Are Not Preventing Contamination from Injected Oil and Gas Wastes*, Report, US General Accounting Office, GAO/RCED-89-9, July (1989).
26. US EPA, *Oil and Gas Injection Wells (Class II)*, US Environmental Protection Agency URL: http://www.epa.gov/safewater/uic/classii.html, Updated October (2004).
27. US EPA, *Mining Wells (Class III)*, US Environmental Protection Agency URL: http://www.epa.gov/safewater/uic/classiii.html, November (2002).
28. US EPA, *Shallow Hazardous and Radioactive Injection Wells (Class IV)*, US Environmental Protection Agency URL: http://www.epa.gov/safewater/uic/classiv.html, Updated October (2008).
29. US EPA, *EPA Proposes to Continue with its Existing Approach for Managing Class V Injection Wells*, US Environmental Protection Agency, EPA 816-F-01-009, April 2001, URL: http://www.epa.gov/safewater/uic/classvdetermination.html, Updated May (2008).
30. US EPA, *The Shallow Injection Wells (Class V)*, US Environmental Protection Agency URL: http://www.epa.gov/safewater/uic/classv.html, August (2008).
31. US EPA, *State of UIC Programs*, US Environmental Protection Agency, URL: http://www.epa.gov/safewater/uic/primacy.html, November (2002).
32. R. P. Selm and B. T. Hulse, Deep well disposal of industrial wastes in *Purdue University Engineering Bulletin*, XLIV, 5, 566 (1959)
33. C. V. Theis, The relation between the lowering of the piezometric surface and the rate and duration of discharge of a well using ground water storage, *Transactions of the American. Geophysical Union*, 16, 519 (1935).
34. US EPA, *Underground Injection Control Program*, Annual Report, US Environmental Protection Agency, Office of Water, December (1991).
35. G. Stewart and W. Pettyjohn, *Development of a Methodology for Regional Evaluation of a Confining Bed Integrity*, EPA/600/2-89/038; US EPA, US Environmental Protection Agency, Office of Research and Development, July (1989).
36. K. E. Davis and R. J. Funk, Experience in deep well disposal of industrial wastes, Water and Wastewater Equipment Manufacturers Association's *Industrial Waste and Pollution Conference and Exposition*, Detroit, MI, April 1–4 (1974).
37. R. M. McKinley, *Temperature, Radioactive Tracer, and Noise Logging for Injection Well Integrity*, EPA/600/R-94/124, US Environmental Protection Agency, Robert S. Kerr Research Center, July (1994).
38. Temple, Barker and Sloane, Inc., *Analysis of Costs Underground Injection Control Regulations Class I, Class III, Class IV and Class V Wells*, EPA Contract No. 68-01-4778.6, May (1979).
39. US ACE, *Civil Works Construction Cost Index System Manual*, US Army Corps of Engineers, Washington, DC, PP 44, (2008-Tables) (2008).
40. D. C. Signor, D. J. Grawitz, and W. Kamn, *Annotated Bibliography on Artificial Recharge of Ground Water, 1955–67*, US Geological Survey, Water Supply Paper 1990 (1970).
41. G. Winqvist and K. Marelius, *The Design of Artificial Recharge Schemes*, England Water Research Association, Vol. I (1970).
42. J, Vocchioli, *Journal of New England Water Works Association*, 136, 2, 87 (1972).
43. L. K. Wang and N. C. Pereira (Eds.), *Water Resources and Natural Control Processes*, Handbook of Environmental Engineering, Vol. 4, p. 228. Humana Press, Clifton, NJ (1986).
44. J. T. Barroclouth, Waste injection into a deep limestone in Northwestern Florida, in *Ground Water*, Vol. 3, Number I (1966).

45. D. Goolsby, Geochemical effects and movements of injected industrial waste in a limestone aquifer, AAPG-USGS *Symposium on Underground Waste Management and Environmental Implications*, Houston, TX (1971).
46. B. J. Schicht, Feasibility of recharging treated sewage effluent into a deep sandstone aquifer, in *National Ground Water Symposium*, Environmental Protection Agency, 16060GR08/71 (1971).
47. US EPA, *Policy Statement on Water Reuse*, Environmental Protection Agency, July 7 (1972).
48. O. J. Esmail and O. K. Kinbler, *Water Resources Research*, 3, 3, 683 (1967).
49. C. W. Sever, *A New and Novel Method for Dewatering, Compacting and Disposing of Sludge*, US Patent Office (1983).
50. D. L. Warner, *Deep-Well Injection in Liquid Waste*, US Department of Health, Education, and Welfare (1965).
51. US EPA, *Report to Congress on Injection of Hazardous Waste*, EPA 57019-85-003, US Environmental Protection Agency, Office of Drinking Water, March (1985).
52. GAO, *Hazardous Waste: Controls Over Injection Well Disposal Operations*, Report to the Chairman, Environment, Energy, and Natural Resources Subcommittee, Committee on Government Operations, House of Representatives, US General Accounting Office, GAO/RCED-87-170, August (1987).
53. Engineering Enterprises, *Assessment of Treatment Technologies Available to Attain Acceptable Levels For Hazardous Waste in Deep Injection Wells*, US Environmental Protection Agency, Underground Injection Control Branch, UIC Contract No. 68-03-3416, October (1987).
54. US EPA, *Analysis of the Effects of EPA Restrictions on the Deep Injection of Hazardous Waste*, EPA 570/9-91-031, US Environmental Protection Agency, Office of Ground Water and Drinking Water, October (1991).
55. A. Apps and C. F. Tsang (eds.), *Deep Injection Disposal of Hazardous and Industrial Waste (Scientific and Engineering Aspects)*, Academic Press, San Diego (1996).
56. US EPA Assessing the Geochemical Fate of Deep-Well-Injected Hazardous Waste, A Reference Guide, EPA 625 6-89 025a, US Environmental Protection Agency, Office of Research and Development, Washington DC (1990).
57. D. L. Warner and J. H. Lehr, *An Introduction to the Technology of Subsurface Waste water Injection*, EPA 600/2-77-240, NTIS PB279 207 (1977).
58. W. F. Langelier, The analytical control of anti-corrosion water treatment, *Journal of American Water Works Association* 28,1500–1 521 (1936).
59. T. E. Larson, and A. M. Buswell, Calcium carbonation saturation index and alkalinity interpretations, *Journal of American Water Works Association* 34,1667–1684 (1942).
60. H. A. Stiff, and L. E. Davis, A method for predicting the tendency of oil field waters to deposit calcium carbonate, *American Institute of Mining Metallurgy, Engineers Transactions, Petroleum Division*, 195, 213–216 (1952).
61. L. K. Wang, Y. T. Hung and N. K. Shammas (eds.) *Advanced Physicochemical Treatment Processes*, Humana Press, Totowa, NJ, (2006).
62. L. K. Wang, Y. T. Hung and N. K. Shammas (eds.) *Physicochemical Treatment Processes*, Humana Press, Totowa, NJ (2005).
63. A. G. Ostroff, *Introduction to Oil Field WaterTechnology*, Prentice-Hall, Englewood Cliffs, NJ (1965).
64. J. W. Watkins, *Analytical Methods of Testing Waters to be Injected into Subsurface Oil-Productive Strata*, US Bureau of Mines Report of Investigations 5031 (1954).
65. J. B. Davis, *Petroleum Microbiology*, Elsevier, New York (1967).

66. J. V. Hunter, Origin of organics from artificial contamination, in *Organic Compounds in Aquatic Environments*, S. D. Faust and J. V. Hunter (eds.), Marcel Dekker, New York, pp. 51–94 (1971).
67. W. R. Roy, S. C. Mravik, I. G. Krapac, D. R. Dickerson, and R. A. Griffin, *Geochemical Interactions of Hazardous Wastes with Geological Formations in Deep-Well Systems*, Environmental Geology Notes (1989).
68. L. R. Reeder, et al., *Review and Assessment of Deep-Well Injection of Hazardous Wastes*, EPA 600/2-77-029a-d, NTIS PB 269 001-004 (1977).
69. A. Strycker, and A. G. Collins, *State-of-the-Art Report: Injection of Hazardous Wastes into DeepWells*, EPA/600/8-87/01 3, NTIS PB87-1 70551 (1987).
70. D. L. Warner and D. H. Orcutt, Industrial wastewater-injection wells in United States-Status of use and regulation, in *Symposium on Underground Waste Management and Artificial Recharge*, J. Braunstein (ed.) Pub. No. 110, International Association of Hydrological Sciences, pp. 687–697 (1973).
71. E. C. Donaldson, *Subsurface Disposal of Industrial Wastes in the United States*, US Bureau of Mines Information Circular 8212 (1964).
72. E. C. Donaldson, R. D. Thomas, and K. H. Johnston, *Subsurface Waste Injections in the United States: Fifteen Case Histories*, US Bureau of Mines Information Circular 8636 (1974).
73. US EPA, *Report to Congress on Injection of Hazardous Wastes* EPA 570/9-85-003 US Environmental Protection Agency, NTIS PB86-203056 (1985).
74. US EPA, *Injection Well Mechanical Integrity Testing*, EPA 625/9-89/0 07, US Environmental Protection Agency (1989).
75. M. Muskat, *Flow of Homogeneous Fluids Through Porous Media*, McGraw-Hill, New York (1937).
76. L. J. Briggs and J. W. McLane, *The Moisture Equivalent of Soils*, US Department of Agriculture, Soils Bull, 45 (1907).
77. L. A. Richards, Capillary conduction of liquids through porous mediums, *Physics*, 1, 318 (1931).
78. US EPA, *Underground Injection Wells and Your Drinking Water*, US Environmental Protection Agency, Office of Water, EPA 813-F-94-001, July (1994).
79. US EPA, *Guide to the Underground Injection Control Program*, US Environmental Protection Agency, June (1979).
80. R. Paul Coneway and Steve Coneway, The T. V. camera see what's wrong, *The Johnson Drillers Journal*, Sept.–Oct. (1978).
81. J. N. Haire and J. D. Haflin, Vertilog-A down-hole casing inspection service, *47th Annual, California Regional Meeting of the Society of Petroleum Engineers of AIME* (1977).
82. M. S. Hantush and C. E. Jacob, *Trans. Am. Geophys. Union*, 36, 95 (1955).
83. D. L. Warner, L. F. Koederitz, A. D. Simon and G. M. You, *Radius of Pressure Influence of Injection Wells*, Environmental Protection Agency, EPA 600/2-79-170, August (1979).
84. US EPA, Part 146 Underground Injection Control Program Criteria and Standards, Environmental Protection Agency, *Fed. Reg*., 45, No 123, 42500, June 24 (1980).
85. L. A. Browning and C. W. Sever, *Simplified Graphic Analysis for Calculation of the Zone of Endangering Influence of an Injection Well*, EPA open files Dallas, TX (1978).
86. L. A. Browning, *Hydraulic Analysis of a Proposed Waste Injection Well, Adams County, Mississippi*, US Environmental Protection Agency open files, Atlanta, GA, March 11 (1980).
87. J. Bear and M. Jacobs, *The Movement of Injected Water Bodies in Confined Aquifers*, Underground Water Storage Study Report No. 13, Techpecon, Haifa, Israel (1964).

88. A. El-Nimir, A. Salih, N. K. Shammas, and M. Khadam, Modeling drained solid matter and salt transport through deep aquifers, *International Workshop on the Application of mathematical Models for Assessment of Changes in Water Quality*, Tunis, Tunisia, 7–12 May (1990).
89. N. K. Shammas, P. Yuan, J. Yang, and Y. T. Hung, Chemical oxidation. In: *Physicochemical Treatment Processes*, L. K Wang, Y. T. Hung, and N. K. Shammas, (eds.), Humana Press, Totowa, NJ, pp. 229–270 (2005).
90. L. K. Wang, D. Vaccari, Y. Li, and N. K. Shammas, Chemical precipitation. In: *Physicochemical Treatment Processes*, L. K Wang, Y. T. Hung, and N. K. Shammas, (eds.), Humana Press, Totowa, NJ, pp. 141–198 (2005).
91. N. K. Shammas, Interactions of temperature, pH and biomass on the nitrification process, *Journal Water Pollution Control Federation*, 58, 1, pp. 52–59, January (1986).
92. N. K. Shammas, An allosteric kinetic model for the nitrification process, *Proceedings of the Tenth Annual Conference of Water Supply Improvement Association*, Honolulu, Hawaii, pp. 1–30, July (1982).
93. L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.) *Biological Treatment Processes*, Humana Press, Totowa, NJ (2009).
94. US EPA, *Class V Injection Wells-Current Inventory, Effects on Ground Water and Technical Recommendations*, Report to Congress, EPA 570/9-87-006, US Environmental Protection Agency, Office of Water, September (1987).
95. US EPA, *Class V Underground Injection Control Study*, Office of Water, EPA/816-R-99-014, 23 volumes and 5 Appendices, September 1999, URL: http://www.epa.gov/safewater/uic/cl5study.html, March (2001).
96. J. B. Foster and D. A. Goolsby, *Construction of Waste-Injection Monitoring Wells near Pensacola, Florida*, Florida Bureau of Geology, Information Circular 74 (1972).
97. J. F. Keely, Chemical time-series sampling, *Ground Water Monitoring Review*, Fall, 29–38 (1982).
98. J. F. Keely and F. Wolf, Field applications of chemical time-series sampling, *Ground Water Monitoring Review* Fall, 26–33 (1983).
99. D. A. Goolsby, Geochemical effects and movement of injected industrial waste in a limestone aquifer. In: *Symposium on Underground Waste Management and Environmental Implications*, T. D. Cook (ed.) American Association of Petroleum Geologists 18, 355–368 (1972).
100. G. Elkan and E. Horvath, *The Role of Microorganisms in the Decomposition of Deep Well Injected Liquid Industrial Waste*, NSF/RA-770102, NTIS PB 268 646 (1977).
101. M. I. Kaufman, D. A. Goolsby, and G. L. Faulkner, Injection of acidic industrial waste into a saline carbonate aquifer: Geochemical aspects. In: *Symposium on Underground Waste Management and Artificial Recharge*, J. Braunstein (ed.) Pub. No. 110, Int. Assn. of Hydrological Sciences, pp. 526–555 (1973).
102. J. A. Leenheer, R. L. Malcolm and W. R. White, *Physical Chemical and Biological Aspects of Subsurface Organic Waste Injection near Wilmington, North Carolina*, US Geological Survey Professional Paper 987 (1 976).
103. J. A. Leenheer, R. L. Malcolm, and W. R. White, Investigation of the reactivity and fate of certain organic compounds of an industrial waste after deep-well injection, *Environmental Science and Technology*, 10, 5, 445–451 (1976).
104. A. DiTommaso and G. H. Elkan, Role of bacteria in decomposition of injected liquid waste at Wilmington, North Carolina. In: *Symposium on Underground Waste Management and Artificial Recharge*, J. Braunstein (ed.), Pub. # 110, International Association of Hydrological Sciences, pp. 585–599 (1973).

APPENDIX

United States Yearly Average Cost Index for Utilities*

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

* Extracted from US ACE (2008-Tables Revised 2008) *Civil Works Construction Cost Index System Manual*, #, US Army Corps of Engineers, Washington, DC (39)

15
Natural Biological Treatment Processes

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract Aquaculture or the production of aquatic organisms (both flora and fauna) under controlled conditions has been practiced for centuries, primarily for the generation of food, fiber, and fertilizer. The water hyacinth and a host of other organisms like duckweed, seaweed, midge larvae, and alligator weeds are used for wastewater treatment. Water hyacinth system, wetland system, evapotranspiration system, rapid rate filtration, slow rate system, overland flow system, and subsurface infiltration have also been applied. This chapter describes the above applications and explains their practice, limitations, design criteria, performance, and costs.

Key Words Aquatic organisms • design and performance • infiltration • natural waste treatment • water hyacinth • weeds • wetland.

1. AQUACULTURE TREATMENT: WATER HYACINTH SYSTEM

1.1. Description

Aquaculture or the production of aquatic organisms (both flora and fauna) under controlled conditions has been practiced for centuries, primarily for the generation of food, fiber and fertilizer. The water hyacinth (*Eichhornia crassipes*) appears to be the most promising organism

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_15 © Humana Press, New York, NY 2009

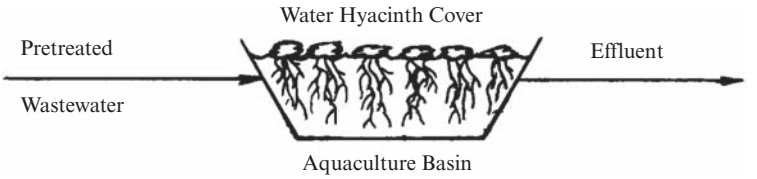


Fig. 15.1. Aquaculture treatment: Water hyacinth system. (*Source*: US EPA (2)).

for wastewater treatment and has received the most attention (1). However, other organisms are being studied. Among them are duckweed, seaweed, midge larvae, alligator weeds and a host of other organisms. Water hyacinths are large fast-growing floating aquatic plants with broad, glossy green leaves and light lavender flowers. A native of South America, water hyacinths are found naturally in waterways, bayous, and other backwaters throughout the South. Insects and disease have little effect on the hyacinth and they thrive in raw, as well as partially treated, wastewater. Wastewater treatment by water hyacinths is accomplished by passing the wastewater through a hyacinth-covered basin (Fig. 15.1), where the plants remove nutrients, BOD_5, suspended solids, metals, etc. Batch treatment and flow-through systems, using single and multiple cell units, are possible. Hyacinths harvested from these systems have been investigated as a fertilizer/soil conditioner after composting, animal feed, and a source of methane when anaerobically digested (2).

1.2. Applications

Water hyacinths are generally used in combination with (following) lagoons, with or without chemical phosphorus removal. A number of full-scale systems are in operation. Most often considered for nutrient removal and additional, treatment of secondary effluent (1–3). Also, research is being conducted on the use of water hyacinths for raw and primary treated wastewater or industrial wastes, but present data favor combination systems. Very good heavy metal uptake by the hyacinth has been reported. Hyacinth treatment may be suitable for seasonal use in treating wastewaters from recreational facilities and those generated from processing of agricultural products. Other organisms and methods with wider climatological applicability are being studied. The ability of hyacinths to remove nitrogen during active growth periods and some phosphorus and retard algae growth provides potential applications in (2, 3):

(a) The upgrading of lagoons.
(b) Renovation of small lakes and reservoirs.
(c) Pretreatment of surface waters used for domestic supply.
(d) Storm water treatment.
(e) Demineralization of water.
(f) Recycling fish culture water.
(g) For biomonitoring purposes.

1.3. Limitations

Climate or climate control is the major limitation. Active growth begins when the water temperature rises above 10 °C. and flourishes when the water temperature is approximately 21 °C. Plants die rapidly when the water temperature approaches the freezing point; therefore, greenhouse structures are necessary in northern locations. Water hyacinths are sensitive to high salinity. Removal of phosphorus and potassium is restricted to the active growth period of the plants.

Metals such as arsenic, chromium, copper, mercury, lead, nickel, and zinc can accumulate in hyacinths and limit their suitability as a fertilizer or feed material. The hyacinths may also create small pools of stagnant surface water, which can breed mosquitoes. Mosquito problems can generally be avoided by maintaining mosquito fish in the system. The spread of the hyacinth plant itself must be controlled by barriers because the plant can spread and grow rapidly and clog affected waterways. Hyacinth treatment may prove impractical for large treatment plants because of land requirements. Removal must be at regular intervals to avoid heavy intertwined growth conditions. Evapotranspiration can be increased by two to seven times greater than evaporation alone.

1.4. Design Criteria

Ponds, channels or basins are in use. In northern climates covers and heat would be required. Harvesting and processing equipment are needed. Operation is by gravity flow and requires no energy. Hyacinth growth energy is supplied by sunlight. All experimental data is from southern climates where no auxiliary heat was needed. Data is not available on heating requirements for northern climates, but it can be assumed proportional to northern latitude of location and to the desired growth rate of hyacinths.

Design data vary widely. Table 15.1 shows the design criteria for water hyacinth systems (4). The following ranges refer to hyacinth treatment as a tertiary process on secondary effluent (2):

(a) Depth should be sufficient to maximize plant rooting and plant absorption.
(b) Detention time depends on effluent requirements and flow, range 4 to 15 days.
(c) Phosphorus reduction 10% to 75%.
(d) Nitrogen reduction 40% to 75%.
(e) Land requirement is usually high 2 to 15 acres/MG/d.

1.5. Performance

Process appears to be reliable from mechanical and process standpoints, subject to temperature constraints. In tests on five different wastewater streams including raw wastewater and secondary effluents, the following removals were reported (2):

(a) BOD_5: 35% to 97%.
(b) TSS: 71% to 83%.
(c) Nitrogen: 44% to 92%.
(d) Total P: 11% to 74%.

Table 15.1
Design criteria for water hyacinth systems

Factor	Aerobic Non aerated	Aerobic Non aerated	Aerobic Aerated
Influent Wastewater	Screened or Settled	Secondary	Screened or Settled
Influent BOD_5, mg/L	130–180	30	130–180
BOD_5 Loading, kg/ha/day	40–80	10–40	150–300
Expected Effluent, mg/L			
BOD_5	<30	<10	<15
SS	<30	<10	<15
TN	< 15	<5	< 15
Water Depth, m	0.6–0.8	0.6–0.9	0.9–1.4
Detention Time, days	10–36	8–18	4–8
Hydsraulic Loading, m^3/ha/day	>200	<800	550–1000
Harvest Schedule	Annualy	Twice per month	Monthly

Source: US EPA (4).

Takeda and Coworkers (3) reported using aquaculture wastewater effluent for strawberry production in a hydroponic system that reduced the final effluent phosphorus concentration to as low as 0.1 mg/L that meets the stringent phosphorus discharge regulations. There is also evidence that in aquaculture system coliform, heavy metals and organics are also reduced, as well as pH neutralization.

Hyacinth harvesting may be continuous or intermittent. Studies indicate that average hyacinth production (including 95% water) is on the order of 1,000 to 10,000 lb/d/acre. Basin cleaning at least once per year results in harvested hyacinths. For further detailed information on water Hyacinth systems the reader is referred to references (5–13).

2. AQUACULTURE TREATMENT: WETLAND SYSTEM

2.1. Description

Aquaculture-wetland systems for wastewater treatment include natural and artificial wetlands as well as other aquatic systems involving the production of algae and higher plants (both submerged and emergent), invertebrates and fish. Natural wetlands, both marine and freshwater, have inadvertently served as natural waste treatment systems for centuries; however, in recent year's marshes, swamps, bogs and other wetland areas have been successfully utilized as managed natural "nutrient sinks" for polishing partially treated effluents under relatively controlled conditions. Constructed wetlands can be designed to meet specific project conditions while providing new wetland areas that also improve available wildlife wetland habitats and the other numerous benefits of wetland areas. Managed plantings of reeds (e.g., *Phragmites* spp.) and rushes (e.g., *Scirpus* spp. and *Schoenoplectus* spp.) as well as managed natural and constructed marshes, swamps, and bogs have been demonstrated to reliably provide pH neutralization and reduction of nutrients, heavy metals, organics, BOD_5, COD, SS, fecal coliforms, and pathogenic bacteria (2, 4).

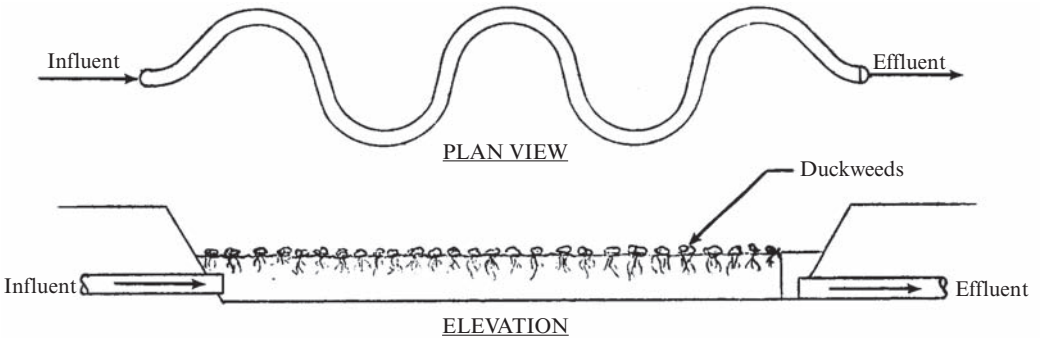


Fig. 15.2. Aquaculture treatment: Wetland system. (*Source*: US EPA (2)).

Wastewater treatment by natural and constructed wetland systems is generally accomplished by sprinkling or flood irrigating the wastewater into the wetland area or by passing the wastewater through a system of shallow ponds, channels, basins or other constructed areas where the emergent aquatic vegetation has been planted or naturally occurs and is actively growing (see Fig. 15.2). The vegetation produced as a result of the system's operation may or may not be removed and can be utilized for various purposes (2):

(a) Composted for use as a source of fertilizer/soil conditioner
(b) Dried or otherwise processed for use as animal feed supplements, or
(c) Digested to produce methane

2.2. *Constructed Wetlands*

Constructed wetlands are classified as a function of water flow (2, 4): surface and subsurface, are known as free water surface (FWS) and subsurface flow system (SFS) [also termed vegetated submerged bed, VSB]. When simply expressed, constructed wetland treatment technology makes artificial receiving water and its vegetation part of the treatment process. In comparison to algae, the higher forms of plant life-floating (duckweed, water hyacinths), submerged, and emergent (cattails, rushes, and reeds) perform less efficiently per unit weight of biomass.

FWS constructed wetland treatment conceptually relies on attached growth bacterial performance, receiving oxygen from the evapotranspiration response of the aquatic vegetation. Practically, the dominant bacterial action is anaerobic. The ammonium and nitrogen removal mechanisms (14–17) are a combination of aerobic oxidation, particulate removal, and synthesis of new plant protoplasm.

An FWS wetland is nothing more than a lagoon, except that a far greater expanse is needed to maximize the productivity per unit area. In practice, very large systems may achieve significant, if not complete, nitrogen oxidation, with surface reaeration contributing to the oxygen supply. Some nitrification and denitrification undoubtedly occurs in all systems.

If it is assumed that the wetland vegetation will not be harvested, as is the case with natural wetland systems, its capacity for nitrogen control is finite, reflecting the site-specific

vegetation and the ability to expand in the available space. Thus, the bigger the natural wetland that is called part of the process, the better, because there is dilution of the wastewater to the point that it is no longer significant in comparison to the naturally occurring background flow and water quality.

Constructed FWS wetlands yield a managed vegetative habitat that becomes an aquaculture system. Examination of the evolution of this technology shows the emergence of concepts that include organic load distribution or artificial aeration to avoid aesthetic nuisances, and emphasis on plants that grow the fastest. Duckweed and water hyacinth systems (classified as aquaculture) have been reported to achieve long-term total nitrogen residuals of less than 10 mg/L and may be manageable, with harvesting and sensitive operation, to values of less than 3 mg/L on a seasonal, if not sustained, basis.

Submerged-flow constructed wetlands are simply horizontal-flow gravel filters with the added component of emergent plants within the media. They have been classically used for BOD removal following sedimentation and/or additional BOD and SS removal from lagoon effluents as with FWS approaches. This technology has the potential for high-level denitrification when a nitrified wastewater is applied; the naturally occurring environment promotes anoxic (denitrification) pathways for oxidized nitrogen elimination.

Ultimately, the success or failure of the wetland approach for nitrogen control may rest with the harvest of the vegetation, the need for backup (so that areas under harvest have the backup of areas in active growth), and often natural seasonal growth and decay cycles. If biomass production is an unacceptable goal, the designer should think of a more tolerant mixed vegetation system that minimizes the need to harvest the accumulated vegetation and maximizes the promotion of concurrent or staged nitrification and denitrification in some fashion. Conceptually, the optimization has to begin with promotion of nitrogen oxidation systems that may be shallow (better aeration for attached and suspended bacterial growth) with vegetation that minimizes light penetration and avoids as much algal growth as possible. Cyclic staging, recycle, forced aeration, and/or mixing represent some of the enhancements that naturally follow (17).

2.3. Applications

Several full-scale systems are in operation or under construction (18). Wetlands are useful for polishing treated effluents. They have potential as a low cost, low energy consuming alternative or addition to conventional treatment systems, especially for smaller flows. Wetlands have been successfully used in combination with chemical addition and overland flow land treatment systems. Wetland systems may also be suitable for seasonal use in treating wastewaters from recreational facilities, some agricultural operations, or other waste-producing units where the necessary land area is available (18). Potential application as an alternative to lengthy outfalls extended into rivers, etc. and as a method of pretreatment of surface waters for domestic supply, storm water treatment, recycling fish culture water and biomonitoring purposes.

2.4. Limitations

Temperature (climate) is a major limitation because effective treatment is linked to the active growth phase of the emergent vegetation. Tie-ins with cooling water from power plants to recover waste heat have potential for extending growing seasons in colder climates. Enclosed and covered systems are possible for very small flows.

Herbicides and other materials toxic to the plants can affect their health and lead to poor treatment. Duckweeds are prized as food for waterfowl and fish and can be seriously depleted by these species. Winds may blow duckweeds to the shore if wind screens or deep trenches are not employed. Small pools of stagnant surface water that can allow mosquitoes to breed can develop, but problems can generally be avoided by maintaining mosquito fish or a healthy mix of aquatic flora and fauna in the system. Wetland systems may prove impractical for large treatment plants because of the large land requirements. They also may cause loss of water because of increases in evapotranspiration.

2.5. Design Criteria

Natural or artificial marshes, swamps, bogs, shallow ponds, channels, or basins could be used. Irrigation, harvesting and processing equipment are optional. Aquatic vegetation is usually locally acquired.

Design criteria are very site and project specific. Available data vary widely. Values below refer to one type of constructed wetland system used as a tertiary process on secondary effluent (2):

(a) Detention Time = 13 days
(b) Land Requirement = 8 acres/MG/d
(c) Depth may vary with type of system, generally 1 to 5 ft

2.6. Performance

Process appears reliable from mechanical and performance standpoints, subject to seasonality of vegetation growth. Low operator attention is required if properly designed.

Tables 15.2 and 15.3 illustrate the capacities of both natural and constructed wetlands for nutrient removal (4). In test units and operating artificial marsh facilities using various wastewater streams, the following removals have been reported for secondary effluent treatment (10 day detention) (2):

(a) BOD_5, 80% to 95%
(b) TSS, 29% to 87%
(c) COD, 43% to 87%
(d) Nitrogen, 42% to 94% depending upon vegetative uptake and frequency of harvesting
(e) Total P, 0% to 94% (high levels possible with warm climates and harvesting)
(f) Coliforms, 86% to 99%
(g) Heavy metals, highly variable depending on species

Table 15.2
Nutrient removal from natural wetlands

Project	Flow, m^3/d	Wetland Type	Percent Reduction			
			TDP[a]	NH_3-N	NO_3-N	TN[b]
Brillion Marsh, WI	757	Marsh	13	–	51	–
Houghton Lake, MI	379	Peatland	95	71	99[c]	–
Wildwood, FL	946	Swamp Marsh	98	–	–	90
Concord, MA	2309	Marsh	47	58	20	–
Bellaire, MI	1136[d]	Peatland	88	–	–	84
Coots Paradise, Town of Dundas, Ontario, Canada	–	Marsh	80	–	–	60–70
Whitney Mobile Park, Home Park, FL	−227	Cypress Dome	91	–	–	89

Source: US EPA (4).
[a] Total dissolved phosphorus.
[b] Total nitrogen.
[c] Nitrate and nitrite.
[d] May-November only.

There is also evidence of reductions in wastewater concentrations of chlorinated organics and pathogens, as well as pH neutralization without causing detectable harm to the wetland ecosystem.

Residuals are dependent upon type of system and whether or not harvesting is employed. Duckweed, for example, yields 50 to 60 lb/acre/d (dry weight) during peak growing period to about half of this figure during colder months. For further detailed information on wetland systems the reader is referred to references (19–23).

3. EVAPOTRANSPIRATION SYSTEM

3.1. Description

Evapotranspiration (ET) system is a means of on-site wastewater disposal that may be utilized in some localities where site conditions preclude soil absorption. Evaporation of moisture from the soil surface and/or transpiration by plants is the mechanism of ultimate disposal. Thus, in areas where the annual evaporation rate equals or exceeds the rate of annual added moisture from rainfall and wastewater application, ET systems can provide a means of liquid disposal without danger of surface or groundwater contamination.

If evaporation is to be continuous, at least three conditions must be met (2):

(a) There must be a continuous supply of heat to meet the latent heat requirement, approximately 590 cal/g of water evaporated at 15 °C.
(b) A vapor pressure gradient must exist between the evaporative surface and the atmosphere to remove vapor by diffusion, convection, or both. Meteorological factors, such as air temperature, humidity, wind velocity, and radiation influence both energy supply and vapor removal.

Table 15.3
Nutrient removal from constructed wetlands

Project	Flow, m^3/day	Wetland Type	BOD_5, mg/L		SS, mg/L		Percent Reduction		Hydraulic surface loading rate, m^3/ha/day
			Influent	Effluent	Influent	Effluent	BOD_5	SS	
Listowol, Ontario (12)	17	FWS[a]	56	10	111	8	82	93	–
Santee, CA (10)	–	SFS[b]	118	30	57	5.5	75	90	–
Sidney, Australia (13)	240	SFS	33	4.6	57	4.5	86	92	–
Arcata, CA	11,350	FWS	36	13	43	31	64	28	907
Emmitsburg, MD	132	SFS	62	18	30	8.3	71	73	1543
Gusline, CA	3,785	FWS	150	24	140	19	84	86	412

Source: US EPA (4).
[a] Free Water Surface System.
[b] Subsurface Flow System.

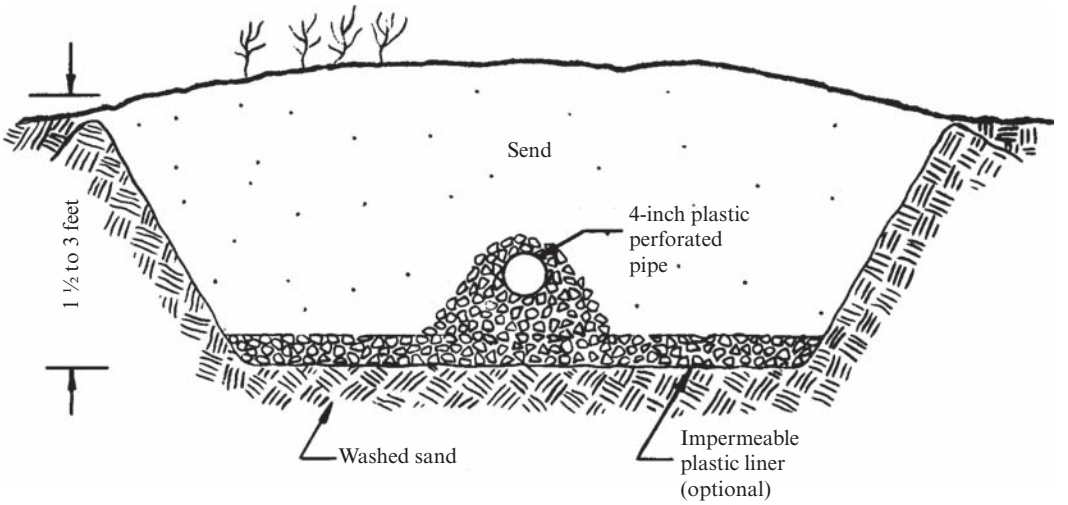


Fig. 15.3. Section through an evapotranspiration bed. (*Source*: US EPA (2)).

(c) There must be a continuous supply of water to the evaporative surface. The soil material must be fine textured enough to draw up the water from the saturated zone to the surface by capillary action but not so fine as to restrict the rate of flow to the surface.

Evapotranspiration is also influenced by vegetation on the disposal field and can theoretically remove significant volumes of effluent in late spring, summer, and early fall, particularly if large silhouette, good transpiring bushes and trees are present.

A typical ET bed system (Fig. 15.3) consists of a 1.5 to 3 ft depth of selected sand over an impermeable plastic liner. A perforated plastic piping system with rock cover is often used to distribute pretreated effluent in the bed. The bed may be square-shaped on relatively flat land, or a series of trenches on slopes. The surface area of the bed must be large enough for sufficient ET to occur to prevent the water level in the bed from rising to the surface.

Beds are usually preceded by septic tanks or aerobic units to provide the necessary pretreatment. Given the proper subsurface conditions, systems can be constructed to perform as both evapotranspiration and absorption beds. Nearly three fourth of all the ET beds in operation were designed to use both disposal methods. Mechanical evaporators have been developed, but are not used at full scale.

3.2. Applications

There are estimated to be 4,000 to 5,000 year-round evapotranspiration beds in operation in the United States, particularly in the semiarid regions of the Southwest.

ET beds are used as an alternative to subsurface disposal in areas where these methods are either undesirable because of groundwater pollution potential or not feasible because of certain geological or physical constraints of land. The ET system can also be designed to supplement soil absorption for sites with slowly permeable soils. The use of ET systems for

summer homes extends the range of application, which is otherwise limited by annual ET rates. Because summer evaporation rates are generally higher and plants with high transpiration rates are in an active growing state, many areas of the country can utilize ET beds for this seasonal application.

3.3. Limitations

The use of an evapotranspiration system is limited by climate and its effect on the local ET rate. In practice, lined ET bed systems are generally limited to areas of the country where pan evaporation exceeds annual rainfall by at least 24 in. The decrease of ET in winter at middle end high latitudes greatly limits its use. Snow cover reflects solar radiation, which reduces EF. In addition, when temperatures are below freezing more heat is required to change frozen water to vapor. When vegetation is dormant, both transpiration and evaporation are reduced. An ET system requires a large amount of land in most regions. Salt accumulation may eventually eliminate vegetation and thus, transpiration. Bed liner (where needed) must be kept water-tight to prevent the possibility of groundwater contamination. Therefore, proper construction methods should be employed to keep the liner from being punctured during installation.

3.4. Design Criteria

Design of an evapotranspiration bed is based on the local annual weather cycle. The total expected inflow based on household wastewater generation and rainfall rates is compared with an average design evaporation value established from the annual pattern. It is recommended to use a 10 year frequency rainfall rate to provide sufficient bed surface area (2). A mass balance is used to establish the storage requirements of the bed. Vegetative cover can substantially increase the ET rate during the summer growing season; but may reduce evaporation during the nongrowing season. Uniform sand in the size range of D_{50} of approximately 0.10 mm is capable of raising water about 3 ft to the top of the bed. The polyethylene liner thickness is typically greater than or equal to 10 mil. Special attention should be paid to storm water drainage to make sure that surface runoff is drained away from the bed proximity by proper lot grading.

3.5. Performance

Performance is a function of climate conditions, volume of wastewater, and physical design of the system. Evapotranspiration is an effective and reliable means of domestic wastewater disposal. An ET system that has been properly designed and constructed is an efficient method for the disposal of pretreated wastewater and requires a minimum of maintenance. Healthy vegetative covers are aesthetically pleasing and the large land requirement, although it limits the land use, it does conserves the open space. Energy is not required, nor is head loss of any value incurred.

3.6. Costs

The following site specific costs serve to illustrate the major components of an evapotranspiration bed in Boulder, Colorado with an annual net ET rate in the range of 0.04 gpd/ft^2 (2).

A 200 gpd household discharge would require a 2 ft deep bed with an area of approximately 5, 000 ft^2. All costs have been adjusted to 2007 US dollars using the Cost Index for Utilities shown in Appendix (24).

Construction cost:

Building sewer with 1,000 gal septic tank, design and permit	$1,633
Excavation and hauling (375 yd^3)	$2,352
Liner (5, 200 ft^2)	$1,524
Distribution piping (625 ft)	$686
Sand (340 yd^3) and gravel (38 yd^3)	$4,094
Supervision and labor	$1,143
Total	$11,432

Annual operation and maintenance cost:

Pumping septage from septic tank (every 3 to 5 years)	$10.89 to $45.73
Total	$10.89 to 45.73

The construction cost for this particular system would be approximately $2.20/$ ft^2, which is consistent with a reported national range of $1.70 to $3.66/$ ft^2. The cost of an evapotranspiration bed is highly dependent upon local material and labor costs. As shown, the cost of sand is a significant portion of the cost of the bed. The restrictive sand size requirement makes availability and cost sensitive to location.

If an aerobic pretreatment unit is used instead of the septic tank add $653 to $6,533 to the construction cost and an amount of $136 to $468/year to the annual operation and maintenance cost.

4. LAND TREATMENT: RAPID RATE SYSTEM

The land-based technologies have been in use since the beginning of civilization. Their greater value may be the use of the wastewater for beneficial return (agricultural and recharge) in water-poor areas, as well as nitrogen control benefits. If nitrogen control benefits are desired, some key issues arise concerning the type of plant crop with its growing and harvesting needs and/or the cycling of the water application and restorative oxygenation resting periods. Native soils and climate add the remaining variables.

Generally, the wastewater applications are cyclic in land-based technologies, making some form of storage or land rotation mandatory to ensure the restorative oxygenation derived from the resting period. Surface wastewater applications allow additional beneficial soil aeration (plowing, tilling, and raking), which can become mandatory for the heavily loaded systems after an elapsed season, or number of loading cycles. Actual surface cleaning programs, to remove the plastic, rubber, and other debris found in pretreated municipal wastewaters, also may be necessary, although not at the frequency used for beneficial soil aeration.

In this and the following sections detailed information on the four most common land-based technologies will be provided. Subsurface, slow, and rapid infiltration systems do not discharge to surface waters and conceptually may allow a more relaxed nitrogen control standard in comparison to the overland flow system, depending on local ground-water regulations.

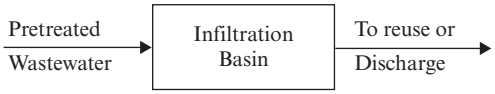


Fig. 15.4. Flow diagram of land treatment using rapid rate system. (*Source*: US EPA (2)).

4.1. Description

Rapid rate infiltration was developed approximately 100 years ago and has remained unaltered since then. It has been widely used for municipal and certain industrial wastewaters throughout the world. Wastewater is applied to deep and permeable deposits such as sand or sandy loam usually by distributing in basins (Fig. 15.4) or infrequently by sprinkling, and is treated as it travels through the soil matrix by filtration, adsorption, ion exchange precipitation, and microbial action (25). Most metals are retained on the soil; many toxic organics are degraded or adsorbed. An underdrainage system consisting of a network of drainage pipe buried below the surface serves to recover the effluent, to control groundwater mounding, or to minimize trespass of wastewater onto adjoining property by horizontal subsurface flow. To recover renovated water for reuse or discharge underdrains are usually intercepted at one end of the field by a ditch. If groundwater is shallow, underdrains are placed at or in the groundwater to remove the appropriate volume of water (2). Thus, the designed soil depth, soil detention time and underground travel distance to achieve the desired water quality can be controlled. Effluent can also be recovered by pumped wells.

Basins or beds are constructed by removing the fine textured top soil from which shallow banks are constructed. The underlying sandy soil serves as the filtration media. Underdrainage is provided by using plastic, concrete (sulfate resistant if necessary), or clay tile lines. The distribution system applies wastewater at a rate that constantly floods the basin throughout the application period of several hours to a couple of weeks. The waste floods the bed and then drains uniformly away, driving air downwards through the soil and drawing fresh air from above. A cycle of flooding and drying maintains the infiltration capacity of the soil material. Infiltration diminishes slowly with time because of clogging. Full infiltration is readily restored by occasional tillage of the surface layer and, when appropriate, removal of several inches from the surface of the basin. Preapplication treatment to remove solids improves distribution system reliability, reduces nuisance conditions, and may reduce clogging rates. Common preapplication treatment practices include the following:

(a) Primary treatment for isolated locations with restricted public access (26).
(b) Biological treatment for urban locations with controlled public access.
(c) Storage is sometimes provided for flow equalization and for nonoperating periods.

Nitrogen removals are improved by (17, 27):

(a) Establishing specific operating procedures to maximize denitrification.
(b) Adjusting application cycles.
(c) Supplying an additional carbon source.
(d) Using vegetated basins (at low rates).

(e) Recycling portions of wastewater containing high nitrate concentrations.
(f) Reducing application rates.

Rapid rate infiltration systems require relatively permeable, sandy to loamy soils. Vegetation is typically not used for nitrogen control purposes but may have value for stabilization and maintenance of percolation rates. The application of algae-laden wastewater to rapid infiltration systems is not recommended because of clogging considerations but could be considered with attendant additional tolerance for surface maintenance, drying and soil aeration needs.

4.2. Applications

Rapid infiltration is a simple wastewater treatment system that is (2):

(a) Less land intensive than other land application systems and provides a means of controlling groundwater levels and lateral subsurface flow.
(b) It provides a means of recovering renovated water for reuse or for discharge to a particular surface water body.
(c) It is suitable for small plants where operator expertise is limited.
(d) It is applicable for primary and secondary effluent and for many types of industrial wastes, including those from breweries, distilleries, paper mills, and wool scouring plants (26, 28, 29).

In very cold weather the ice layer floats atop the effluent and also protects the soil surface from freezing. Generated residuals may require occasional removals of top layer of soil. The collected material is disposed of onsite.

4.3. Limitations

The rapid infiltration process is limited by (2):

(a) Soil type
(b) Soil depth
(c) The hydraulic capacity of the soil
(d) The underlying geology, and
(e) The slope of the land

Nitrate and nitrite removals are low unless special management practices are used.

4.4. Design Criteria

The design criteria for rapid rate system can be summarized as follows (2):

(a) Field area 3 to 56 acres/MG/d
(b) Application rate 20 to 400 ft/year, 4 to 92 in./week
(c) BOD_5 loading rate 20 to 100 lb/acre/d
(d) Soil depth 10 to 15 ft or more
(e) Soil permeability 0.6 in./h or more
(f) Hydraulic loading cycle 9 hours to 2 weeks application period, 15 hours to 2 weeks resting period
(g) Soil texture sands, sandy barns
(h) Basin size 1 to 10 acres, at least 2 basins/site
(i) Height of dikes 4 ft; underdrains 6 ft or more deep

Table 15.4
Loading cycles for high rate infiltration systems

Loading cycle objective	Applied wastewater	Season	Application period, d[a]	Drying period, d
Maximize infiltration rates	Primary	Summer	1–2	5–7
		Winter	1–2	7–12
	Secondary	Summer	1–3	4–5
		Winter	1–3	5–10
Maximize nitrogen removal	Primary	Summer	1–2	10–14
		Winter	1–2	12–16
	Secondary	Summer	7–9	10–15
		Winter	9–12	12–16
Maximize nitrification	Primary	Summer	1–2	5–7
		Winter	1–2	7–12
	Secondary	Summer	1–3	4–5
		Winter	1–3	5–10

Source: US EPA (25).
[a] Regardless of season or cycle objective, application periods for primary effluent should be limited to 1–2 days to prevent excessive soil clogging.

(j) Application techniques: flooding or sprinkling
(k) Preapplication treatment: primary or secondary.

Designs can be developed that foster only nitrification or nitrification and denitrification (17, 27). Nitrification is promoted by low hydraulic loadings and short application periods (1 to 2 days) followed by long drying periods (10 to 16 days). Denitrification can vary from 0% to 80%. For significant denitrification, the application period must be long enough to ensure depletion of the soil (and nitrate nitrogen) oxygen. Higher denitrification values predictably track higher BOD: nitrogen ratios. Enhancement may be promoted by recycling or by adding an external driving substrate (methanol). Nitrogen elimination strategies also may reduce the drying period by about half to yield lower overall nitrogen residuals with higher ammonium-nitrogen concentrations. Suggested loading cycles (25) to maximize infiltration rates, nitrogen removal and nitrification rates are given in Table 15.4.

4.5. Performance

The effluent quality is generally excellent where sufficient soil depth exists and is not normally dependent on the quality of wastewater applied within limits. Well designed systems provide for high quality effluent that may meet or exceed primary drinking water standards. Percent removals for typical pollution parameters are (2):

(a) BOD_5, 95% to 99 %
(b) TSS, 95% to 99%
(c) Total N, 25% to 90%
(d) Total P, 0% to 90%t until flooding exceeds adsorptive capacity (30)
(e) Fecal Coliform, 99.9 to 99.99 + % (31)

The process is extremely reliable, as long as sufficient resting periods are provided. However, it has a potential for contamination of groundwater by nitrates. Heavy metals could be eliminated by pretreatment techniques as necessary. Monitoring for metals and toxic organics is needed where they are not removed by pretreatment. The process requires long term commitment of relatively large land areas, although small by comparison to other land treatment systems (32, 33).

4.6. Costs

The construction and operation & maintenance costs are shown in Figs. 15.5 and 15.6 respectively (2). The costs are based on 1973 (Utilities Index = 149.36, US EPA Index 194.2, ENR Index = 1, 850) figures. To obtain the values in terms of the present 2007 US dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of 3.61 (24).

Assumptions applied in preparing the costs given in Fig. 15.5 and Fig. 15.6:

(a) Application rate 182 ft/year.
(b) Construction costs include field preparations (removal of brush and trees) for multiple unit infiltration basins with 4 ft dike formed from native excavated material, and storage is not assumed necessary.
(c) Drain pipes buried 6 to 8 ft with 400 ft spacing, interception ditch along length of field, and weir for control of discharge; gravel service roads and 4-ft stock fence around perimeter.

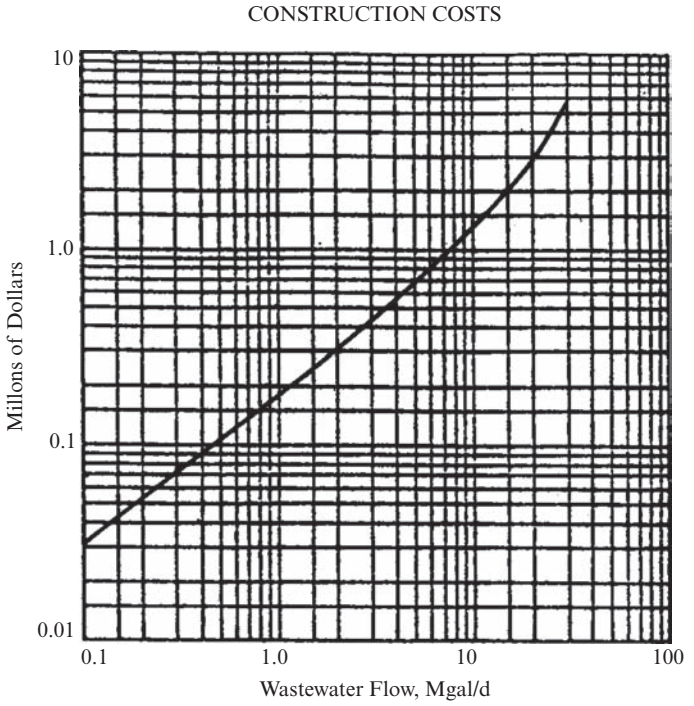


Fig. 15.5. Construction costs for rapid rate system. (*Source*: US EPA (2)).

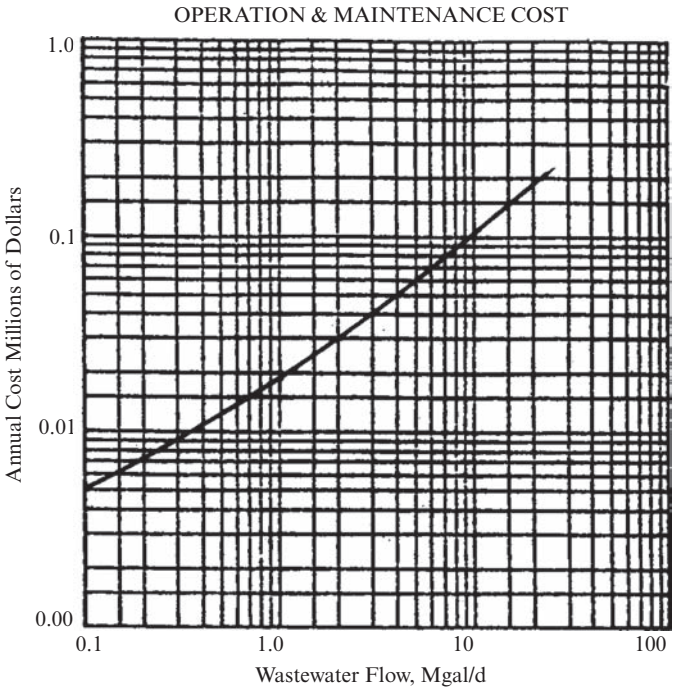


Fig. 15.6. Operation and maintenance costs of rapid rate system. (*Source*: US EPA (2)).

(d) O & M cost includes inspection and unclogging of drain pipes at outlets; annual tilling of infiltration surface and major repair of dikes after 10 years; high pressure jet cleaning of drain pipes every 5 years, annual cleaning of interceptor ditch, and major repair of ditches, fences and roads after 10 years.

(e) Costs of pretreatment monitoring wells, land and transmission to and from pretreatment facility not included.

5. LAND TREATMENT: SLOW RATE SYSTEM

5.1. Description

Slow rate land treatment represents the predominant municipal land treatment practice in the United States. In this process, wastewater is applied by sprinkling to vegetated soils that are slow to moderate in permeability (clay barns to sandy barns) and is treated as it travels through the soil matrix by filtration, adsorption, ion exchange, precipitation, microbial action and by plant uptake (Fig. 15.7). An underdrainage system consisting of a network of drainage pipe buried below the surface serves to recover the effluent, to control groundwater, or to minimize trespass of leachate onto adjoining property by horizontal subsurface flow. To recover renovated water for reuse or discharge, underdrains are usually intercepted at one end of the field by a ditch. Underdrainage for groundwater control is installed as needed to prevent waterlogging of the application site or to recover the renovated water for reuse. Proper crop

Fig. 15.7. Flow diagram of land treatment using slow rate system. (*Source*: US EPA (2)).

management also depends on the drainage conditions. Sprinklers can be categorized as hand moved, mechanically moved, and permanent set, the selection of which includes the following considerations (2):

(a) Field conditions (shape, slope, vegetation, and soil type)
(b) Climate
(c) Operating conditions
(d) Economics

Vegetation is a vital part of the process and serves to extract nutrients, reduce erosion and maintain soil permeability. Considerations for crop selection include:

(a) Suitability to local climate and soil conditions.
(b) Consumptive water use and water tolerance.
(c) Nutrient uptake and sensitivity to wastewater constituents.
(d) Economic value and marketability.
(e) Length of growing season.
(f) Ease of management.
(g) Public health regulations.

Common preapplication treatment practices include the following:

(a) Primary treatment for isolated locations with restricted public access and when limited to crops not for direct human consumption.
(b) Biological treatment plus control of coliform to 1000 MPN/100 mL for agricultural irrigation, except for human food crops to be eaten raw.
(c) Secondary treatment plus disinfection to 200 MPN/100 mL fecal coliform for public access areas (parks).

Wastewaters high in metal content should be pretreated to avoid plant and soil contamination. Table 15.5 shows the wastewater constituents that have potential adverse effects on crops (25). Forestland irrigation is more suited to cold weather operation, because soil temperatures are generally higher, but nutrient removal capabilities are less than for most field crops.

5.2. Applications

Slow rate systems produce the best results of all the land treatment systems. Advantages of sprinkler application over gravity methods include (34):

(a) More uniform distribution of water and greater flexibility in range of application rates.
(b) Applicability to most crops.
(c) Less susceptibility to topographic constraints.
(d) Reduced operator skill and experience requirements.

Table 15.5
Potential adverse effects of wastewater constituents on crops

Problem and related constituent	Constituent level			Crops affected
	No problem	Increasing problems	Severe problems	
Salinity (BC_W), mmho/cm	<0.75	0.75–3.0	>3.0	Crops in arid climates only (see Table 9-4)
Specific ion toxicity from root absorption				
Boron, mg/L	<0.5	0.5–2	2.0–10.0	Fruit and citrus trees – 0.5–1.0 mg/L; field crops – 1.0–2.0 mg/L; grasses – 2.0–10.0 mg/L
Sodium, adj–SAR[a]	<3	3.0–9.0	>9.0	Tree crops
Chloride, mg/L	<142	142–355	>355	Tree crops
Specific ion toxicity from foliar absorption				
Sodium, mg/L	<69	>69	–	Field and vegetable crops under sprinkler application
Chloride, mg/L	<106	>106	–	
Miscellaneous				
$NH_4 - N + NO_3 - N$, mg/L	<5	5–30	30	Sugarbeets, potatoes, cotton, grains
HCO_3, mg/L	<90	90–520	>520	Fruit
pH, units	6.5–8.4	4.2–5.5	<4.2 and >8.5	Most crops

Source: US EPA (25).
[a] Adjusted sodium adsorption ratio.

Underdrainage provides a means of recovering renovated water for reuse or for discharge to a particular surface water body when dictated by senior water rights and a means of controlling groundwater. The system also provides the following benefits:

(a) An economic return from the use of water and nutrients to produce marketable crops for forage.
(b) Water and nutrient conservation when utilized for irrigating landscaped areas.

5.3. Limitations

The slow rate process is limited by (2):

(a) Soil type and depth
(b) Topography
(c) Underlying geology
(d) Climate
(e) Surface and groundwater hydrology and quality
(f) Crop selection

(g) Land availability

Crop water tolerances, nutrient requirements, and the nitrogen removal capacity of the soil-vegetation complex limit hydraulic loading rate (35). Climate affects growing season and will dictate the period of application and the storage requirements. Application ceases during period of frozen soil conditions. Once in operation, infiltration rates can be reduced by sealing of the soil. Limitations to sprinkling include adverse wind conditions and clogging of nozzles. Slopes should be less than 15% to minimize runoff and erosion. Pretreatment for removal of solids and oil and grease serves to maintain reliability of sprinklers and to reduce clogging. Many states have regulations regarding preapplication disinfection, minimum buffer areas, and control of public access for sprinkler systems.

The process requires long term commitment of large land area; i.e., largest land requirement of all land treatment processes (36). Concerns with aerosol carriage of pathogens, potential vector problems, and crop contamination have been identified, but are generally controllable by proper design and management.

5.4. Design Criteria

The design criteria for slow rate system can be summarized as follows (2):

(a) Field area 56 to 560 acres/MG/d
(b) Application rate 2 to 20 ft/year, 0.5 to 4 in./week
(c) BOD_5 loading rate 0.2 to 5 lb/acre/d
(d) Soil depth 2 to 5 ft or more
(e) Soil permeability 0.06 to 2.0 in./h
(f) Minimum preapplication treatment primary
(g) Lower temperature limit 25 °F
(h) Particle size of solids less than 1/3 sprinkler nozzle diameter
(i) Underdrains 4 to 8 in. diameter, 4 to 10 ft deep, 50 to 500 ft apart, pipe material plastic, concrete (sulfate-resistant, if necessary) or clay.

5.5. Performance

Effluent quality is generally excellent and consistent regardless of the quality of wastewater applied (37). Percent removals for typical pollution parameters when wastewater is applied through more than 5 ft of unsaturated soil are:

(a) BOD_5, 90% to 99 + %
(b) TSS, 90% to 99 + %
(c) Total N, 50% to 95% depending on N uptake of vegetation
(d) Total P, 80% to 99%, until adsorptive capacity is exceeded (38)
(e) Fecal Coliform, 99.99 + % when applied levels are more than 10 MPN/100 mL

This treatment is capable of achieving the highest degree of nitrogen removal. Typically, nitrogen losses because of denitrification (15% to 25%), ammonia volatilization (0% to 10%) and soil immobilization (0% to 25%) supplement the primary nitrogen removal mechanism by the crop (17). The balance of the nitrogen passes to the percolate. Typical design standards require preservation of controlling depths to ground water and establishing nitrogen limits

in either the percolate or ground water as it leaves the property site. Nitrogen loading to the ground water is often the controlling consideration in the design. For further detailed information on slow rate infiltration systems the reader is referred to references (39–44).

5.6. Costs

The construction and operation & maintenance costs are shown in Figs. 15.8 and 15.9 respectively (2). The costs are based on 1973 (Utilities Index = 149.36, US EPA Index 194.2, ENR Index = 1850) figures. To obtain the values in terms of the present 2007 US dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of 3.61 (24).

Assumptions applied in preparing the costs given in Figs. 15.8 and 15.9:

(a) Yearly average application rate: 0.33 in./d
(b) Energy requirements: Solid set spray distribution requires 2,100 kWh/year/ft of TDH/MG/d capacity. Center pivot spraying requires an additional 0.84 × 106 kWh/year/acre (based on 3.5 d/week operation) for 1 MG/d or larger facilities (below 1 MG/d, additional power = 0.84 to 1.35 × 106 kWh/year/acre)
(c) Clearing costs are for brush with few trees using bulldozer-type equipment
(d) Solid set spraying construction costs include: lateral spacing, 100 ft; sprinkler spacing, 80 ft along laterals; 5.4 sprinklers/acre; application rate, 0.20 in./h; 16.5 gpm flow to sprinklers at 70 psi; flow to laterals controlled by hydraulically operated automatic valves; laterals buried 18 in.; mainlines buried 36 in.; all pipe 4 in. diameter and smaller is PVC; all larger pipe is asbestos cement (Total dynamic head = 150 ft)
(e) Center pivot spraying construction costs include: heavy-duty center pivot rig with electric drive; multiple units for field areas over 40 acres; maximum area per unit, 132 acres; distribution pipe is buried 3 ft deep

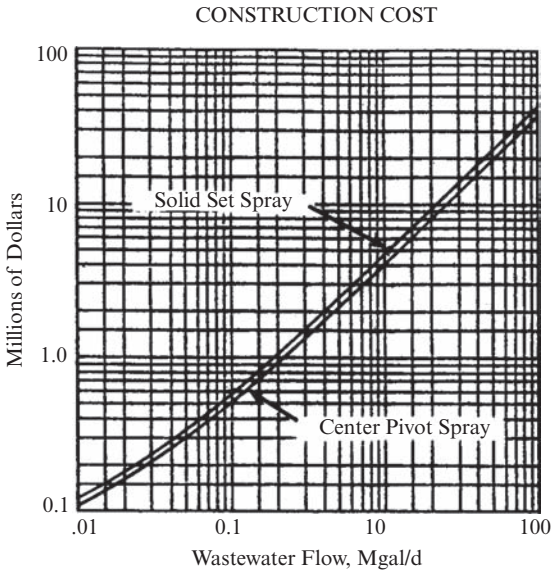


Fig. 15.8. Construction cost of slow rate system. (*Source*: US EPA (2)).

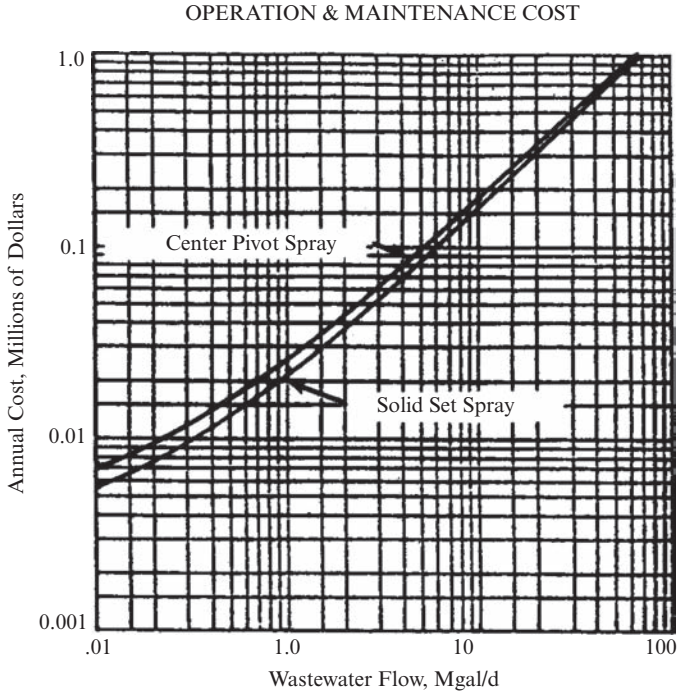


Fig. 15.9. Operation and maintenance cost of slow rate system. (*Source*: US EPA (2)).

(f) Underdrains are spaced 250 ft between drain pipes. Drain pipes are buried 6 to 8 ft deep with interception ditch along length of field and weir for control of discharge
(g) Distribution pumping construction costs include: structure built into dike of storage reservoir; continuously cleaned water screens; pumping equipment with normal standby facilities; piping and valves within structure; controls and electrical work
(h) Labor costs include inspection and unclogging of drain pipes at outlets and dike maintenance
(i) Materials costs include for solid set spraying: replacement of sprinklers and air compressors for valve controls after 10 years; for center pivot spraying, minor repair parts and major overhaul of center pivot rigs after 10 years; high pressure jet cleaning of drain pipes every 5 years, annual cleaning of interceptor ditch, and major repair of ditches after 10 years; distribution pumping repair work performed by outside contractor and replacement parts; scraping and patching of storage receiver liner every 10 years
(j) Storage for 75 days is included; 15 ft dikes (12-ft wide at crest) are formed from native materials (inside slope 3:1, outside 2:1); rectangular shape on level ground; 12-ft water depth; multiple cells for more than 50 acre size; asphaltic lining; 9-in. riprap on inside slope of dikes
(k) Cost of pretreatment, monitoring wells, land, and transmission to and from land treatment facility not included.

6. LAND TREATMENT: OVERLAND FLOW SYSTEM

6.1. Description

Wastewater treatment using the overland flow system is relatively new. It is now extensively used in the food processing industry. Very few municipal plants are in operation and most are in warm, dry areas. A flow diagram of the system is shown in Fig. 15.10. Wastewater is applied over the upper reaches of sloped terraces and is treated as it flows across the vegetated surface to runoff collection ditches. The wastewater is renovated by physical, chemical and biological means as it flows in a thin film down the relatively impermeable slope.

A secondary objective of the system is for crop production. Perennial grasses (Reed Canary, Bermuda, Red Top, tall fescue and Italian Rye) with long growing seasons, high moisture tolerance and extensive root formation are best suited to overland flow. Harvested grass is suitable for cattle feed. Biological oxidation, sedimentation and grass filtration are the primary removal mechanisms for organics and suspended solids. Nitrogen removal is attributed primarily to nitrification/denitrification and plant uptake. Loading rates and cycles are designed to maintain active microorganism growth on the soil surface. The operating principles are similar to a conventional trickling filter with intermittent dosing. The rate and length of application is controlled to minimize severe anaerobic conditions that result from overstressing the system. The resting period should be long enough to prevent surface ponding, yet short enough to keep the microorganisms in an active state. Surface methods of distribution include the use of gated pipe or bubbling orifice. Gated surface pipe, which is attached to aluminum hydrants, is aluminum pipe with multiple outlets. Control of flow is accomplished with slide gates or screw adjustable orifices at each outlet. Bubbling orifices are small diameter outlets from laterals used to introduce flow. Gravel may be necessary to dissipate energy and ensure uniform distribution of water from these surface methods. Slopes must be steep enough to prevent ponding of the runoff, yet mild enough to prevent erosion and provide sufficient detention time for the wastewater on the slopes. Slopes must have a uniform cross slope and be free from gullies to prevent channeling and allow uniform distribution over the surface. The network of slopes and terraces that make up an overland system may be adapted to natural rolling terrain. The use of this type of terrain will minimize land preparation costs. Storage must be provided for nonoperating periods. Runoff is collected in open ditches. When unstable soil conditions are encountered or flow velocities are erosive, gravity pipe collection systems may be required. Common preapplication practices include the following:

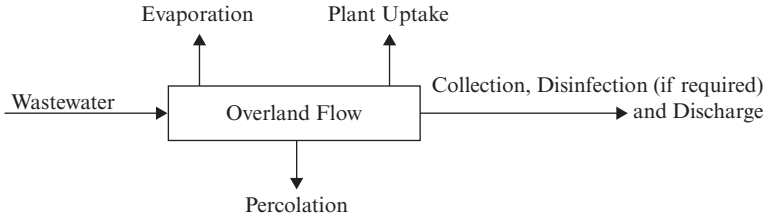


Fig. 15.10. Flow diagram of land treatment using overland flow system. (*Source*: US EPA (2)).

screening or comminution for isolated sites with no public access; screening or comminution plus aeration to control odors during storage or application for urban locations with no public access (45, 46). Wastewaters high in metal content should be pretreated to avoid soil and plant contamination.

A common method of distribution is with sprinklers. Recirculation of collected effluent is sometimes provided and/or required. Secondary treatment before overland flow permits reduced (as much as two third reduction) land requirements. Effluent disinfection is required where stringent fecal coliform criteria exist.

6.2. Application

Because overland flow is basically a surface phenomenon, soil clogging is not a problem. High BOD_5 and suspended solids removals have been achieved with the application of raw comminuted municipal wastewater. Thus, preapplication treatment is not a prerequisite where other limitations are not operative. Depth to groundwater is less critical than with other land systems. It also provides the following benefits: an economic return from the reuse of water and nutrients to produce marketable crops or forage; and a means of recovering renovated water for reuse or discharge. This type of applications is preferred for gently sloping terrain with impermeable soils.

6.3. Limitations

The process is limited by soil type, crop water tolerances, climate, and slope of the land. Steep slopes reduce travel time over the treatment area and thus, treatment efficiency. Flat land may require extensive earthwork to create slopes. Ideally, slope should be 2% to 8%. High flotation tires are required for equipment. Cost and impact of the earthwork required to obtain terraced slopes can be major constraints. Application is restricted during rainy periods and stopped during very cold weather (47). Many states have regulations regarding preapplication disinfection, minimum buffer zones and control of public access.

6.4. Design Criteria

The design criteria for overland Flow system can be summarized as follows (2):

(a) Field area required, 35 to 100 acres/MG/d.
(b) Terraced slopes 2% to 8%.
(c) Application rate, 11 to 32 ft/year, 2.5 to 16 in./week.
(d) BOD_5 loading rate 5 to 50 lb/acre/d.
(e) Soil depth, sufficient to form slopes that are uniform and to maintain a vegetative cover.
(f) Soil permeability 0.2 in./h or less.
(g) Hydraulic loading cycle 6 to 8 hours application period, 16 to 181 wresting period.
(h) Operating period 5 to 6 d/week.
(i) Soil texture clay and clay loams.

Below are representative application rates for 2% to 8% sloped terraces:

In./week	Pretreatment	Terrace length, ft
2.5 to 8	untreated or primary	150
6 to 16	Lagoon or secondary	120

Table 15.6
Design loadings for overland flow systems

Preapplication treatment	Application rate $m^3/h\ m$	Hydraulic loading rate cm/d
Screening/Primary	0.07–0.12[a]	2.0–7.0[b]
Aerated Cell (1 day detention)	0.08–0.14	2.0–8.5
Wastewater Treatment pond[c]	0.09–0.15	2.5–9.0
Secondary[d]	0.11–0.17	3.0–10.0

Source: US EPA (48).
[a] $m^3/h\ m \times 80.5 = gal/h\ ft.$
[b] $cm/d \times 0.394 = in./d.$
[c] Does not include removal of algae.
[d] Recommended only for upgrading existing secondary treatment.

Generally, 40% to 80% of applied wastewater reaches collection structures, lower percent in summer and higher in winter (southwest data). Table 15.6 shows the required pretreatment and allowed application and hydraulic rates (48).

6.5. Performance

Percent removals for comminuted or screened municipal wastewater over about 150 ft of 2% to 6% slope:

(a) BOD_5, 80% to 95%
(b) Suspended solids, 80% to 95%
(c) Total N, 75% to 90%
(d) Total P, 30% to 60%,
(e) Fecal coliform 90% to 99.9%

The addition of alum [$Al_2\ (SO_4)_3$], ferric chloride [$FeC1_3$] or calcium carbonate [$CaCO_3$] before application will increase phosphorus removals.

Little attempt has been made to design optimized overland flow systems with a specific objective of nitrogen control. Their performance depends on the same fundamental issues: nitrification-denitrification, ammonia volatilization, and harvesting of crops. When measured, overland flow systems designed for secondary treatment often reveal less than 10 mg/L total nitrogen (49). For further detailed information on overland flow systems the reader is referred to references (50–53).

6.6. Costs

The construction and operation & maintenance costs are shown in Figs. 15.11 and 15.12 respectively (2). The costs are based on 1973 (Utilities Index = 149.36, US EPA Index 194.2, ENR Index = 1, 850) figures. To obtain the values in terms of the present 2007 US dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of 3.61 (24).

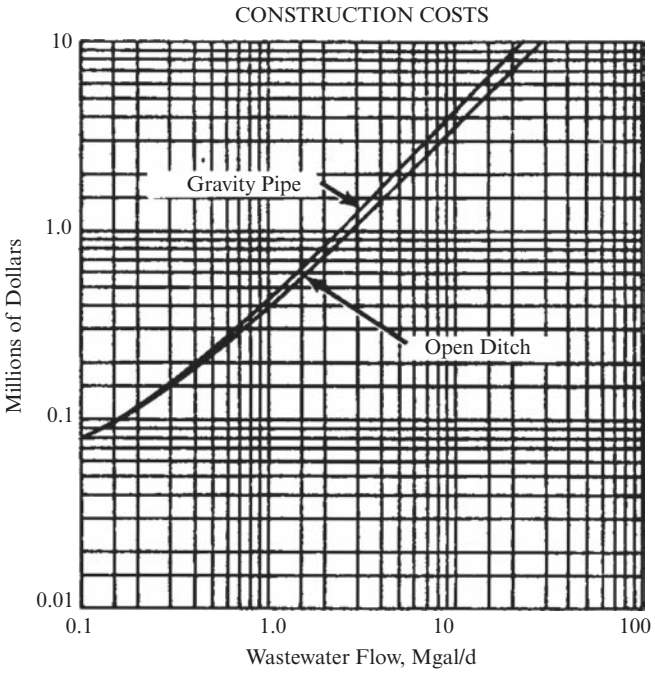


Fig. 15.11. Construction cost of overland flow treatment system. (*Source*: US EPA (2)).

Assumptions applied in preparing the costs given in Figs. 15.11 and 15.12:

(a) Storage for 75 days included.
(b) Site cleared of brush and trees using bulldozer-type equipment; terrace construction: 175 to 250 ft wide with 2.5% slope (1,400 yd/acre of cut). Costs include surveying, earthmoving, finish grading, ripping two ways, disking, land-planning and equipment mobilization.
(c) Distribution system: application rate, 0.064 in./h; yearly average rate of 3 in./week (8 h/d; 6 d/week); flow to sprinklers, 13 gpm at 50 psi; laterals 70 ft from top of terrace, buried 18 in; flow to laterals controlled by hydraulically operated automatic valves; mainlines buried 36 in; all pipe 4 in diameter and smaller is PVC: all larger pipe is asbestos cement.
(d) Open Ditch Collection: network of unlined interception ditches sized for a 2 in./h storm; culverts under service roads; concrete drop structures at 1,000 ft intervals.
(e) Gravity Pipe Collection: network of gravity pipe interceptors with inlet/manholes every 250 ft along sub-mains; storm runoff is allowed to pond at inlets; each inlet/manhole serves 1,000 ft of collection ditch; manholes every 500 ft along interceptor mains.
(f) O & M cost includes replacement of sprinklers and air compressors for valve controls after 10 years and either biannual cleaning of open ditches with major repair after 10 years or the periodic cleaning of inlets and normal maintenance of gravity pipe. Also includes dike maintenance and scraping and patching of storage basin liner every 10 years.
(g) Costs for pretreatment, land, transmission to site, disinfection and service roads and fencing not included.

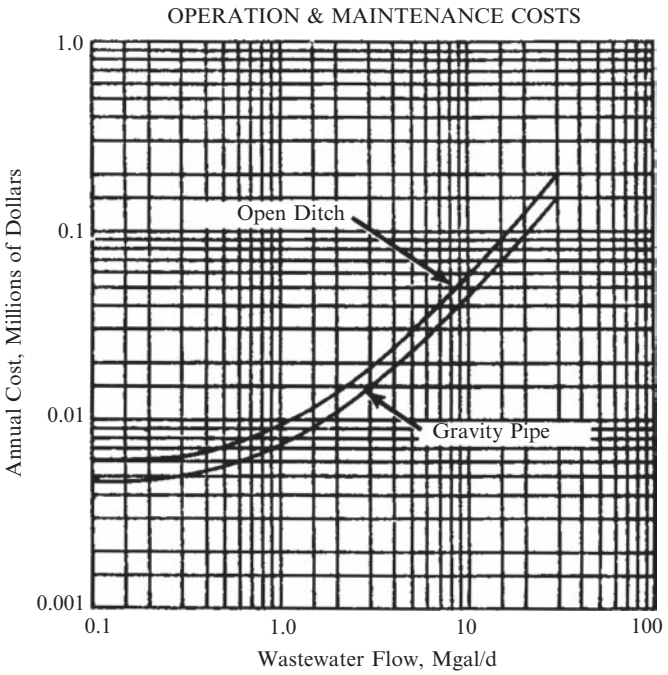


Fig. 15.12. Operation and maintenance cost of overland flow treatment system. (*Source*: US EPA (2)).

7. SUBSURFACE INFILTRATION

Subsurface infiltration systems are capable of producing a high degree of treatment; with proper design, they can provide a nitrified effluent, and denitrification can be achieved under certain circumstances. Keys to their success are the adequacy of the initial gravel infiltration zone for solids capture and the following unsaturated zone of native or foreign soils. Failure to provide an oxygenated environment by either resting or conservative loadings can lead to failure. Denitrification under gravity loading is likely to be small, but may be improved through pressure/gravity dosing concepts of liquid application to the trenches (54).

Subsurface infiltration wastewater management practices are embodied in the horizontal leach fields that routinely serve almost one-third of the United States population that use more than 20 million septic tanks in their individual nonsewered establishments and homes (2). In recent years, they have also been advanced for collective service in small isolated communities.

7.1. Description

A septic tank followed by a soil absorption field is the traditional on-site system for the treatment and disposal of domestic wastewater from individual households or establishments. The system consists of a buried tank where wastewater is collected and scum, grease, and settleable solids are removed by gravity separation, and a sub-surface drainage system where

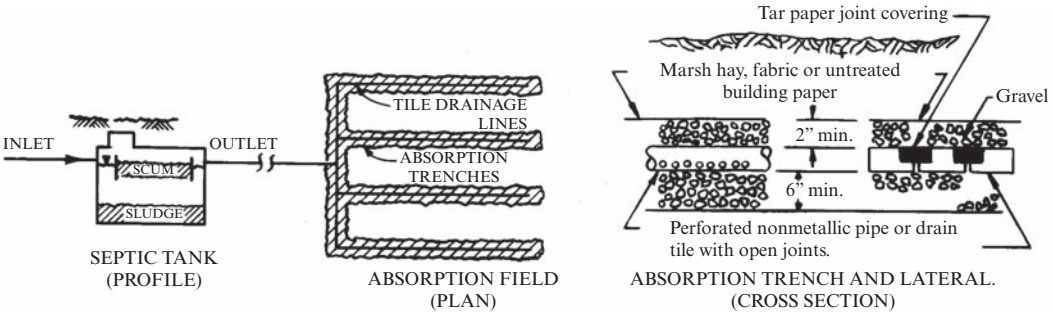


Fig. 15.13. Septic tank absorption field. (*Source*: US EPA (2)).

clarified effluent percolates into the soil. Precast concrete tanks with a capacity of 1000 gal are commonly used for house systems. Solids are collected and stored in the tank, forming sludge and scum layers. Anaerobic digestion occurs in these layers, reducing the overall volume. Effluent is discharged from the tank to one of three basic types of subsurface systems, absorption field (54), seepage bed (54, 55), or seepage pits (56). Sizes are usually determined by percolation rates, soil characteristics, and site size and location. Distribution pipes are laid in a field of absorption trenches to leach tank effluent over a large area (Fig. 15.13). Required absorption areas are dictated by state and local codes. Trench depth is commonly about 24 in. to provide minimum gravel depth and earth cover. Clean, graded gravel or similar aggregate, varying in size from ½ to 2½ in., should surround the distribution pipe and extend at least two inches above and six inches below the pipe. The maintenance of at least a 2 ft separation between the bottom of the trench and the high water table is required to minimize groundwater contamination. Piping typically consists of agricultural drain tile, vitrified clay sewer pipe, or perforated, nonmetallic pipe. Absorption systems having trenches wider than 3 ft are referred to as seepage beds. Given the appropriate soil conditions (sandy soils), a wide bed makes more efficient use of available land than a series of long, narrow trenches.

Many different designs may be used in laying out a subsurface disposal field. In sloping areas, serial distribution can be employed with absorption trenches by arranging the system so that each trench is utilized to its capacity before liquid flows into the succeeding trench. A dosing tank can be used to obtain proper wastewater distribution throughout the disposal area and give the absorption field a chance to rest or dry out between dosings. Providing two separate alternating beds is another method used to restore the infiltrative capacity of a system. Aerobic units may be substituted for septic tanks with no changes in soil absorption system requirements.

In areas where problem soil conditions preclude the use of subsurface trenches or seepage beds, mounds can be installed (Fig. 15.14) to raise the absorption field above ground, provide treatment, and distribute the wastewater to the underlying soil over a wide area in a uniform manner (2, 57, 58). A pressure distribution network should be used for uniform application of clarified tank effluent to the mound. A subsurface chamber can be installed with a pump

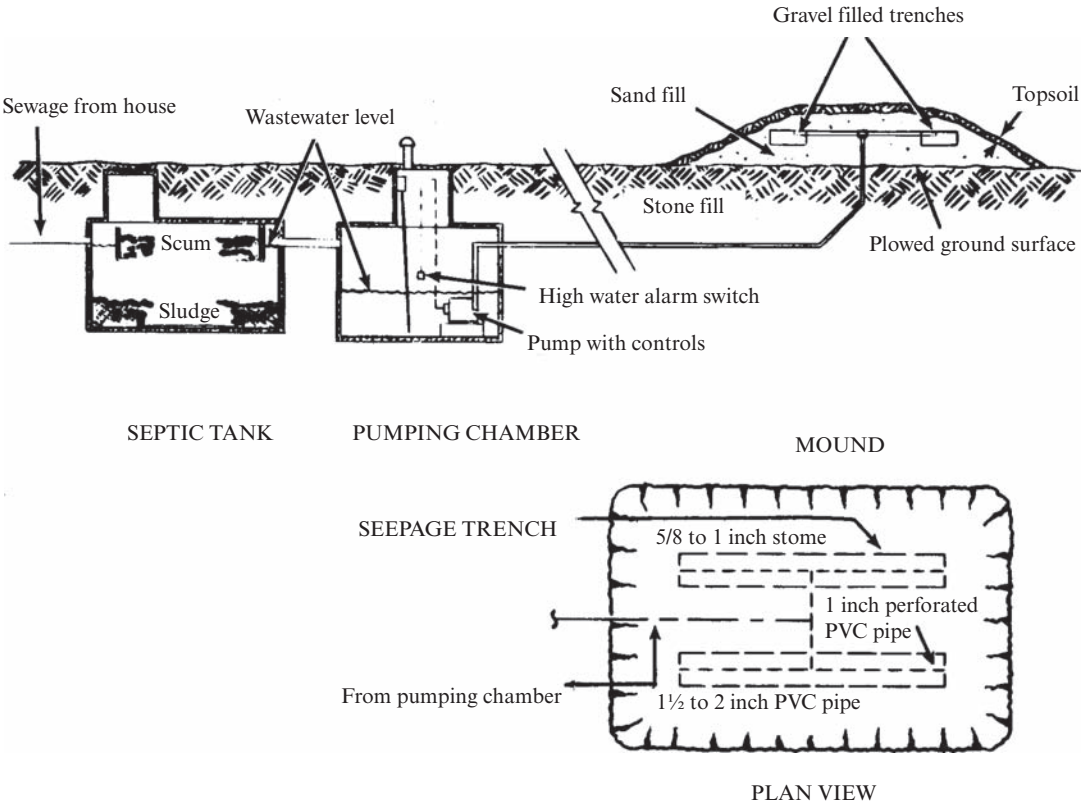


Fig. 15.14. Figure 15.13 septic tank mound absorption field. (*Source*: US EPA (2)).

and high water alarm to dose the mound through a series of perforated pipes. Where sufficient head is available, a dosing siphon may be used. The mound must provide an adequate amount of unsaturated soil and spread septic tank effluent over a wide enough area so that distribution and purification can be effected before the water table is reached.

The mound system requires more space and periodic maintenance than conventional subsurface disposal system, along with higher construction costs. System cannot be installed on steep slopes, nor over highly (120 min/in.) impermeable subsurface. Seasonal high groundwater must be deeper than two feet to prevent surfacing at the edge of the mound (2). An alternative to the mound system is a new combined distribution and pretreatment unit to precede the wastewater application to the subsurface infiltration systems (59). The new system is based on pumping of septic tank effluent to one or more units filled with lightweight clay aggregates. The wastewater is distributed evenly over the 2.3 m^2 surface of the pretreatment filter. The filter(s) effluent is then applied to the subsurface infiltration system.

7.2. Applications

Subsurface infiltration systems for the disposal of septic tanks effluents are used primarily in rural and suburban areas where economics are favorable. Properly designed and installed systems require a minimum of maintenance and can operate in all climates.

7.3. Limitations

The use of subsurface effluent disposal fields is dependent on the following factors and conditions (2):

(a) Soil and site conditions.
(b) The ability of the soil to absorb liquid.
(c) Depth to groundwater.
(d) Nature of and depth to bedrock.
(e) Seasonal flooding.
(f) Distance to well or surface water.

A percolation rate of 60 min/in. is often used as the lower limit of permeability. The limiting value for seasonal high groundwater should be 2 ft below the bottom of the absorption field. When a soil system loses its capacity to absorb septic tank effluent, there is a potential for effluent surfacing, which often results in odors and, possibly, health hazards.

7.4. Design Criteria

Absorption area requirements for individual residences are given in Table 15.7. The area required per bedroom is a function of the percolation rate, the higher the rate the smaller is the required area (2).

Design criteria for the mound system is as follows (2, 57, 58): Design flow 75 gal/person/day; 150 gal/bedroom/day. Basal area based on percolation rates up to

Table 15.7
Required areas of subsurface infiltration absorption fields

Percolation. rate, min/in.	Required area per bedroom, ft^2
1 or less	70
3	100
5	125
10	165
15	190
30	250
45	300
60	330

Source: US EPA (2).

120 mm/in. Mound height at center approximately 3.5 to 5 ft. Pump (centrifugal) must accommodate approximately 30 gpm at required TDH.

Properly designed, constructed, and operated septic tank systems have demonstrated an efficient and economical alternative to public sewer systems, particularly in rural and sparsely developed areas. System life for properly sited, designed, installed and maintained systems may equal or exceed 20 years.

7.5. Performance

Performance is a function of the following factors (2):

(a) Design of the system components.
(b) Construction techniques employed.
(c) Rate of hydraulic loading.
(d) Area geology and topography.
(e) Physical and chemical composition of the soil mantle.
(f) Care given to periodic maintenance.

Pollutants are removed from the effluent by natural adsorption and biological processes in the soil zone adjacent to the field. BOD, SS, bacteria, and viruses, along with heavy metals and complex organic compounds, are adsorbed by soil under proper conditions. However, chlorides and nitrates may readily penetrate coarser, aerated soils to groundwater.

Leachate can contaminate groundwater when pollutants are not effectively removed by the soil system. In many well aerated soils, significant densities of homes with septic tank-soil absorption systems have resulted in increasing nitrate content of the ground water. Soil clogging may result in surface ponding with potential aesthetic and public health problems. The sludge and scum layers accumulated in a septic tank must be removed every 3 to 5 years.. For further detailed information on subsurface infiltration systems the reader is referred to references (60–66).

Additional technical information on the emerging natural biological treatment processes can be found from the literature (67–71).

REFERENCES

1. FINS Information Service, *Constructed Wetland for Aquaculture Wastewater*, Aquatic Technology, http://fins.actwin.com/aquatic-plants/month.9612/msg00372.htlm Dec. (1996).
2. US EPA, Innovative and alternative Technology Assessment Manual, EPA/430/9-78-009, US Environmental Protection Agency, Washington, DC, (1980).
3. F. Takeda, P. R. Adler, and D. M. Glen, Strawberry production linked to aquaculture wastewater treatment, International Society for Horticultural Science, *ISHS III International Strawberry Symposium*, Veldhoven, Netherland 1 September (1997). Abstract can be located at www.actahort.org/books/439/439_113.htm.
4. US EPA, *Design Manual: Constructed Wetlands and Aquatic Plant Systems for Municipal Wastewater Treatment*, EPA/625/1-88/022, US Environmental Protection Agency, Office of Research and Development, Washington, DC, September (1988).
5. Metcalf and Eddy, *Wastewater Engineering Treatment and Reuse*, 4th ed., McGraw Hill, New York (2003).

6. A. Vesilind, *Wastewater Treatment Plant Design*, Water Environment Federation and IWA Publishing, Alexandria, VA, USA (2003).
7. L. K. Wang, and N. C. Pereira, (eds.), *Handbook of Environmental Engineering, Vol. 3, Biological Treatment Processes*, Humana Press, New Jersey, pp. 520 (1986).
8. Microtack, *Organic Aquaculture and Wastewater Treatment Supplies*, TechOzone, Bangok, Thailand, www.microtack.com (2002).
9. W. F. DeBusk and K. R. Reedy, Wastewater treatment using floating aquatic macrophytes: contaminant removal processes and management strategies. In: *Aquatic Plants for Water Treatment and Resource Recovery*, K. R. Reedy and W. H. Smith (eds.), Magnolia Publishing, pp. 27–48 (1987).
10. K. R. Reedy and D. L. Sutton, Water hyacinth for water quality improvement and biomass production, *Journal of Environmental Quality*, 14, 459–462 (1984).
11. G. Tchobanoglous, F. Maitski, K. Thomson and T. H. Chadwick, Evolution and performance of City of San Diego pilot scale aquatic Wastewater Treatment system using water hyacinth, *J. Water Pollut. Control Fed*., 61, 11/12 (1989).
12. S. C. Reed and R. W. Crites, *Handbook of Land Treatment Systems for Industrial and Municipal Wastes*, Noyes Publications, Park Ridge, NJ (1984).
13. S. C. Reed, R. W. Crites and E. J. Middlebrooks, *Natural Systems for Waste Management and Treatment*, McGraw-Hill, NY (1995).
14. N. K. Shammas, *Optimization of Biological Nitrification*, Ph.D. dissertation, Microfilm Publication, University of Michigan, Ann Arbor, Michigan (1971).
15. L. K. Wang, Chemistry of nitrification-denitrification process, *Journal of Environmental Science*, 21, 23–28, (1978).
16. L. K. Wang and D. B. Aulenbach, *BOD and Nutrient Removal by Biological A/O Process Systems*, US Depart. Of Commerce, National Technical information Service Technical Report #PB88-168430/AS, pp.12, December (1986).
17. US EPA, *Manual Nitrogen Control*, EPA/625/R-93/010, US Environmental Protection Agency, Office of Research and Development, Washington, DC, September (1993).
18. Y. T. Hung, S. Gubba, H. Lo, L. K. Wang, C. Yapijakis, and N. K. Shammas, Application of wetland for wastewater treatment, *OCEESA Journal*, 20, 1, 41–46, (2003).
19. R. W. Crites, Constructed wetlands for wastewater treatment and reuse, presented at the *Engineering Foundation Conference, Environmental and Engineering Food Processing Industries XXVI*, Santa Fe, New Mexico (1996).
20. WPCF, Natural Systems for Wastewater Treatment, Manual of Practice # FD-16, Water Pollution Control Federation, Alexandria, VA, February (1990).
21. D. A Hammer (ed.), Constructed Wetlands for Wastewater Treatment; municipal, Industrial and Agricultural, Lewis Publishers, Chelsea, MI (1989).
22. J. C. Wang, D. B. Aulenbach, L. K Wang, and M. H. S. Wang, Energy models and cost models for water pollution controls, Chapter 29. In: *Clean Production*, K. B. Misra (ed.), Springer-Verlag, Berlin, Germany, pp.685–720 (1996).
23. L. K. Wang, J. V. Krougzek, and U. Kounitson, *Case Studies of Cleaner Production and Site Rededication*, United Nations Industrial Development Organization (UNIDO), Vienna, Austria, UNIDO-Registry No. DTT-5-4-95, 136 pp., April (1995).
24. US ACE, *Civil Works Construction Cost Index System Manual*, US Army Corps of Engineers, Washington, DC, (2008-Tables Revised) (2008).
25. US EPA, Process *Design Manual, Land Treatment of Municipal Wastewater*, EPA 625/1-81-013, US Environmental Protection Agency, Center for Environmental Research Information, Cincinnati, OH (1981).

26. M. B. Satterwhite, B. J. Condike and G. L. Stewart, *Treatment of Primary Sewage Effluent by Rapid Infiltration*, US Army Corps of Engineers, Cold Region Research and Engineering Laboratory, December (1976).
27. R. W. Crites, Nitrogen removal in rapid infiltration system, *Journal of Environmental Engineering, ASCE*, 111, 865 (1985).
28. D. G. Smith, K. D. Linstedt and E. R. Bennett, *Treatment of Secondary Effluent by Infiltration-percolation*, U. S. Environmental Protection Agency, EPA-600/2-79-174, August (1979).
29. H. Bouwer, Renovating secondary effluent by groundwater Recharge with Infiltration Basins, *Conference on Recycling Treated Municipal wastewater Through Forest and Cropland*, U. S. Environmental Protection Agency, EPA-660/2-74-003 (1974).
30. D. R. Kioussis, F. W. Wheaton, and P. Kofinas Phosphate binding polymeric hydrogels for aquaculture wastewater remediation, *Aquaculture Engineering*, 19, 3, 163–178, Feb. (1999).
31. C. P. Gerba and J. C. Lance, Pathogen removal from wastewater during ground water recharge, *proceedings of Symposium on Wastewater Reuse for Groundwater Recharge*, Pomona, CA, Sept. 6–7 (1979).
32. D. B. Aulenbach, *Long Term Recharge of Trickling Filter Effluent into Sand*, US Environmental Protection Agency, EPA-600/2-79-068, March (1979).
33. E. Leach, C. G. Enfield, and C. C. Harlin, Jr., *Summary of Long-Term Rapid Infiltration System Studies*, US Environmental Protection Agency, EPA-600/2-80-165, July (1980).
34. US EPA, Evaluation of Land Application Systems, US Environmental Protection Agency, EPA-430/9-75-001, March (1975).
35. N. K. Shammas, Investigation of irrigation water application rates to landscaped areas in Ar-Riyadh, *Journal of Engineering Sciences*, 3, 2, 147–165 (1991).
36. R. Stone and J. Rowlands, Long-Term effects of Land Application of Domestic Wastewater: Mesa, Arizona Irrigation Site, US Environmental Protection Agency, EPA-600/2-80-061, April (1980).
37. A. Uiga and R. W. Crites, Relative health risks of activated sludge treatment and slow rate land treatment, *Journal of Water Pollution Control Federation*, 52, 12, 2865–2874 (1980).
38. T. J. Tofflemire and M. Chen, Phosphate removal by sands and soils. In: *Land as a Waste Management Alternative*, R. C. Loehr (ed.), Ann Arbor Science, Ann Arbor, MI (1977).
39. US Department of Agriculture, Soil Conservation Service, Trickling Irrigation, Chapter 7. In: *Irrigation, SCS National Engineering Handbook*, U. S. Government Printing Office, Washington, DC (1986).
40. T. F. Jenkins and E. J. Palazzo, Wastewater Treatment by a Slow Rate Land Treatment System, US Army Corps of Engineers, Cold Region Research and Engineering Laboratory, CRREL Report 81-14, Hanover, NH, August (1981).
41. R. C. Loehr (ed.), *Land as a Waste Management Alternative*, Ann Arbor Science, Ann Arbor, MI (1977).
42. A. R. Overman, Wastewater Irrigation at Tallahassee, Florida, US Environmental Protection Agency, EPA-600/2-79-151, August (1979).
43. L. A. Duscha, Dual Cropping Procedure for Slow Infiltration of Land Treatment of Municipal Wastewater, Department of the Army, Engineering Technical Letter # 1110-2-260, March (1981).
44. N. K. Shammas and A. El-Rehaili, Tertiary filtration of wastewater for use in irrigation", *Symposium on the Effect of Water Quality on the Human Health and Agriculture*, Al-Khobar, Saudi Arabia, October (1986).

45. R. G. Smith and E. D. Schroeder, Field studies of the overland flow Process for the treatment of raw and primary treated municipal wastewater, *Journal of Water Pollution Control Federation*, 57, 7 (1985).
46. L. E. Perry, E. J. Reap and M. Gilliand, Evaluation of the overland flow process for the treatment of high-strength food processing wastewaters, *Proceedings of the 14th Mid-Atlantic Industrial Waste Conference*, University of Maryland, June (1982).
47. R. F. de Figueredo, R. G. Smith, and E. D. Schroeder, Rainfall and overland flow performance, *Journal of Environmental Engineering, ASCE*, 110, 678 (1984).
48. US EPA, Process Design Manual, Land Treatment of Municipal Wastewater; Supplement on Rapid Infiltration and Overland Flow, EPA 625-1-81-13a, US Environmental Protection Agency Center.
49. J. Johnston, R. F. Smith, and E. D. Schroeder, Operating schedule effects on nitrogen removal in overland flow wastewater treatment systems, presented at the *61st Annual Water Pollution Control Federation Conference*, Dallas, TX (1988).
50. J. L. Witherow and B. E. Bledsoe, Algae removal by the overland flow process, *Journal of Water Pollution Control Federation*, 55, 1256 (1983).
51. S. C. Reed and R. W. Crites, *Handbook of Land Treatment Systems for Industrial and Municipal Wastes*, Noyes Publications, Park Ridge, NJ (1984).
52. R. G. Smith and E. D. Schroeder, Physical design of overland flow systems, *Journal of Water Pollution Control Federation*, 55, 3 (1983).
53. L. K Wang, Wastewater treatment by biological physicochemical two-stage process system, *Proceedings of the 41st Annual Purdue Industrial Waste Conference*, Lafayette, IN, p. 67 (1987).
54. R. J. Otis, G. D. Plews, and D. H. Patterson, Design of conventional soil absorption trenches and beds, ASAE, *Proceedings of the 2nd National Home Sewage Treatment Symposium*, Chicago, IL, December (1977).
55. T. W. Bendixen, J. B. Coulter, and G. M. Edwards, *Study of Seepage Beds*, Robert A. Taft Sanitary Engineering Center, Cincinnati, OH (1960).
56. T. W. Bendixen, R. E. Thomas, and J. B. Coulter, Report of a Study to Develop Practical Design Criteria for Seepage Pits as a method for Disposal of Septic Tank Effluents, NTIS Report # PB 216 931, Cincinnati, OH, pp. 252 (1963).
57. J. C. Converse, B. L. Carlile and G. B. Peterson, Mounds for the treatment and disposal of septic tank effluent, ASAE, Proceedings of the 2nd National Home Sewage Treatment Symposium, Chicago, IL December (1977).
58. J. C. Converse, Design and Construction Manual for Wisconsin Mounds, Small Scale Waste Management Project, University of Wisconsin, Madison, WI, pp. 80 (1978).
59. ASAE, *A New Combined Distribution and Pretreatment Unit for Wastewater Soil Infiltration Systems*, American Society of Agricultural Engineer, Technical Library, http://asae.frymulti.com, (2004)
60. F. W. Krof, R. Laak and K. A. Healey, Equilibrium operation of subsurface Absorption Systems, *Journal of Water Pollution Control Federation*, 49, 2007–2016 (1977).
61. Absorption Systems, *Journal of Water Pollution Control Federation*, 49, 2007–2016 (1977).
62. W. L Mellen, Identification of Soils as a Tool for the Design of Individual Sewage Disposal Systems, Lake County Health Department, Waukegan, IL, p. 67 (1976).
63. A.P. Bernhardt, Treatment and Disposal of Wastewater from Homes by Soil Infiltration and Evapotranspiration, University of Toronto Press, Toronto, Canada, p. 173 (1978).
64. A. P. Bernhart, Return of effluent nutrients to the natural cycle through evapotranspiration and subsoil-infiltration of domestic wastewater, ASCE, *Proceedings of the National Home Sewage Disposal Symposium*, Chicago, IL, December (1974).

65. NEHA, *On-Site Wastewater Management*, National Environmental Health Association, Denver, CO, p. 108 (1979).
66. US EPA, Design Manual: *Onsite Wastewater Treatment and Disposal Systems*, EPA 625/1-80-012, US Environmental Protection Agency, Office of Research and Development, Municipal Environmental Research Laboratory, Cincinnati, OH, October (1980).
67. M. Krofta, and L. K. Wang, *Treatment of Household Wastes, Septage and Septic Tank Effluent, Technical Report PB-2000-101750INZ*, U. S. Dept. Of Commerce, National Technical Information Service, Springfiled, VA (1999).
68. K. D. White and L. K. Wang, Natural treatment and on-site processes, *Water Environment Research*, 72, 5, 1–12 (2000).
69. D. B. Aulenbach and N. L. Clesceri, Treatment by application onto land. In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. H. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ (2009).
70. N. L. Clesceri, D. B. Aulenbach, and J. F. Roetzer, Treatment by surface application. In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ (2009).
71. L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.), *Biosolids Treatment Processes*, Humana Press, Totowa, NJ, 820 pp. (2007).

APPENDIX

United States yearly average Cost Index for Utilities (24)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

16
Emerging Suspended-Growth Biological Processes

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract Among the emerging suspended-growth biological treatment processes covered in this chapter are powdered activated carbon treatment (PACT) process, carrier-activated sludge process (CAPTOR and CAST systems), activated bio-filter (ABF), vertical loop reactor (VLR), and phostrip process. This chapter describes the above processes and explains their practice, limitations, design criteria, energy requirements, process equipment, performance, and costs.

Key Words ABF•CAPTOR•CAST•PACT•suspended-growth•VLR and phostrip processes.

1. POWDERED ACTIVATED CARBON TREATMENT (PACT)

1.1. Types of PACT Systems

The powdered activated carbon (PAC) activated sludge system is a process modification of the activated sludge process. PAC is added to the aeration tank where it is mixed with the biological solids (Fig. 16.1). The mixed liquor solids are settled and separated from the treated effluent. In a gravity clarifier, polyelectrolyte will normally be added before the clarification step to enhance solid-liquid separation. If phosphorus removal is necessary, alum is often added at this point also. Even with polyelectrolyte addition, tertiary filtration is normally required to reduce the level of effluent suspended solids. The clarifier underflow solids are

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_16

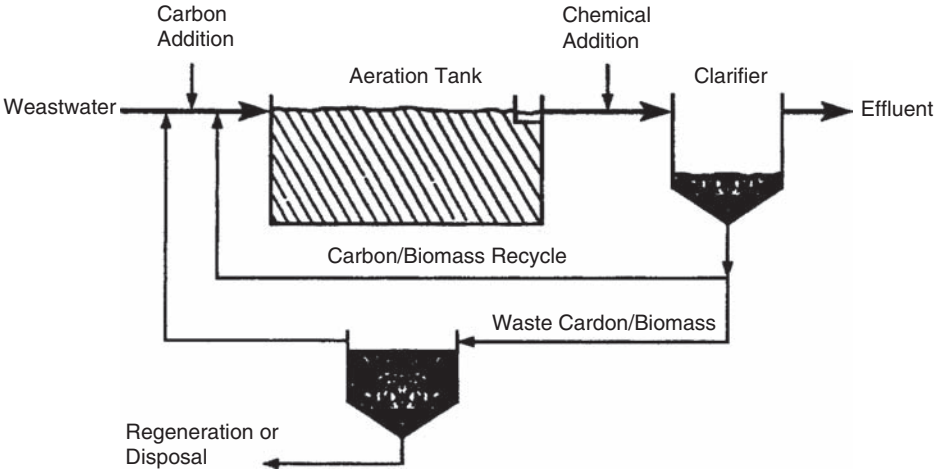


Fig. 16.1. Powdered activated carbon activated sludge process (PACT) (10, 14).

continuously returned to the aeration tank. A portion of the carbon-biomass mixture is wasted periodically to maintain the desired solids inventory in the system.

There are six types of combined biological and physicochemical PAC process systems (1–7):

(a) Continuous combined biological and physicochemical PAC process systems involving the use of sedimentation clarifiers.
(b) Combined biological and physicochemical PAC sequencing batch reactor systems involving the use of sedimentation clarifiers.
(c) Continuous combined biological and physicochemical PAC process systems involving the use of dissolved air flotation (DAF) clarifiers.
(d) Combined biological and physicochemical PAC sequencing batch reactor systems involving the use of DAF clarifiers.
(e) Continuous combined biological and physicochemical PAC process systems involving the use of membrane filters (MF).
(f) Combined biological and physicochemical PAC sequencing batch reactor involving the use of membrane filters (MF).

When PAC is dosed into an activated sludge process for combined adsorption and biochemical reactions, the combined process is also called PACT process, in which PAC still stands for powdered activated carbon, whereas ACT stands for activated sludge.

1.2. Applications and Performance

The addition of PAC to plug flow and complete mix suspended growth reactors is a more common process modification for industrial wastewater treatment than for municipal systems. Demonstrated advantages of PAC addition to suspended growth reactors include (8):

(a) Improved solids settling and dewatering characteristics.
(b) The ability of PAC to adsorb biorefractory materials and inhibitory compounds.

(c) Improving effluent quality and reducing the impact of organic shock loads.
(d) Reduction in odor, foaming, and sludge bulking.
(e) Improved color and 5-day BOD removal.

Because PAC is wasted with excess biomass, virgin or regenerated PAC addition is required to maintain the desired concentration in the biological reactor. This can represent a significant cost factor for the system. When carbon addition requirements exceed 900 to 1,800 kg/day (2,400 to 4,000 lb/day), wet air oxidation/regeneration (WAR) is claimed to represent an economical approach to carbon recovery and waste biomass destruction (9). However, an ash separation step is needed in this case, affecting the economics of carbon regeneration and recovery (10). The economic analysis is further clouded by the inability to analytically differentiate powdered carbon from background refractory volatile materials, thus making it difficult to quantify the value of the volatile suspended material recovered after WAR. Although ash separation processes have been reported to be effective in at least two municipal PAC activated sludge plants, the economics of complete PAC/WAR systems relative to other activated sludge nitrification systems are unclear (7, 10, 11).

In the United States, PACT systems for nitrification generally have been applied at municipal treatment plants where industrial sources contribute a significant fraction of the incoming wastewater. In all instances PAC regeneration was included in the flowsheet (12). A summary of selected municipal PACT facilities is presented in Table 16.1.

The procedure to follow in designing PACT systems for nitrification involves a modification to those for complete mix or conventional plug flow systems to account for the effects of the addition of PAC (13). According to the major supplier of the technology (12, 14), most

Table 16.1
Summary of PACT Process Systems using wet air oxidation for APC regeneration (10, 14)

Facility	Current/Design Flow, m^3/s	PAC/WAR[a] Status	Reason for PAC[b]	Permit Limits		
				BOD_5, mg/L	TSS, mg/L	NH_4^+-N, mg/L
Vemon, CT	0.18/.28	MA	C	10	20	–
Mt. Holly, NJ	0.11/.22	MA	C,S	30	30	20
E. Burlington, NC	0.31/.53	MA	C,N,T	12–24	30	4.0–8.0
S. Burlington, NC	0.30/.42	AS	C,N,T	12–24	30	4.0–8.0
Kalamazoo, MI	1.1/2.4	MA	C,N,T	7–30	20–30	2.0–10.0
Bedford Hts., OH	0.15/.15	NAC	N,S	10	12	5.1
Medina Co., OH	0.31/.44	MA	N	10	12	1.5–8.0
N. Oimsted,[c] OH	0.26/.31	AS	N,S	30	30	2.3–8.9
Sauget, IL	0.70/1.2	AS	T	20	25	–
EI Paso, TX	0.20/.44	MA	N,O	–	–	–

[a] MA = Modified operation and/or design for ash control AS = Converted to conventional activated sludge. NAC = Converted to the use of nonactivated carbon without regeneration.
[b] C = Color Removal; S = Space; N = Nitrification; T = Toxics; O = Organics.
[c] Plan to convert to NAC without regeneration.

PAC process systems are designed at MLSS concentrations of approximately 15 g/L The mixed liquor is composed of volatile activated carbon, biomass, nonvolatile PAC ash, biomass decay components, and influent inert material. The relative proportions of these materials are strongly influenced by whether carbon regeneration via wet air oxidation and a return of this material to the aerator is practiced. The intent is to maintain the PAC concentration at approximately 1.5 times the biomass level in nitrification PAC reactors (12, 14). The most appropriate PAC concentration will be dictated by the specific wastewater characteristics and often cannot be specified without bench or pilot scale studies. The PAC concentration to be added will depend on the design solids retention time, the hydraulic retention time and the required PAC concentration in the reactor. According to the U.S. Environmental Protection Agency (14), for practical engineering design considering the loss, the PAC concentration to be added can be calculated from Eq. (1):

$$\mathrm{PACI} = \mathrm{PACE} + (\mathrm{PACR})\,\mathrm{HRT}/\mathrm{SRT} \tag{1}$$

where:

PACI = influent PAC concentration, mg/L
PACR = mixed liquor PAC concentration in the reactor, mg/L
PACE = effluent PAC concentration, mg/L
HRT = hydraulic retention time, day
SRT = design solids retention time, day

The value of PACE in Eq. (1) can be estimated by assuming that the carbon fraction in the effluent TSS (total suspended solids) is the same as the fraction of PAC in the MLSS (mixed liquor suspended solids).

PACT nitrification systems are normally selected when the municipal wastewater contains compounds originating from industrial operations, as stated previously. Nitrifiers are susceptible to a number of organic and inorganic inhibitors found in many industrial wastewaters (14). Researchers have provided evidence that the addition of PAC to nitrifying activated sludge systems receiving industrial wastewaters improved nitrification rates (14–16). More recent studies have been completed with the goal of determining the mechanism of nitrification enhancement in PAC activated sludge systems in the presence of adsorbable and nonadsorbable inhibitors (17). The results indicated that the addition of the proper amount of PAC can completely nullify the toxic effects of an adsorbable nitrification inhibitor. A minor positive effect on nitrification rates was observed when PAC was added to a nitrifying activated sludge system receiving nonadsorbable inhibitors. The activated sludge used in these studies was not acclimated to the inhibiting compounds. Another possible contributing factor to the enhancement of nitrification could be attributed to the fact that the addition of PAC provides particulate matter for attachment of the nitrifying microorganisms, thereby promoting nitrification (18).

1.3. Process Equipment

PAC can be fed in the dry state using volumetric or gravimetric feeders or can be fed in slurry form. There are more than three major PAC producers, over 50 manufacturers of volumetric and gravimetric feeders, and over 50 manufacturers of slurry feeders (19–21). There are also many manufacturers of sequencing batch reactors (SBR) (2), dissolved air flotation (DAF) clarifiers (7), and membrane filtration (MF) reactors (6).

1.4. Process Limitations

The process limitations of PACT process systems are identical to that of the PAC physicochemical process. PACT process will increase the amount of generated sludge. Regeneration will be necessary at higher dosages to maintain reasonable costs. Most systems will require post-filtration to capture any residual carbon particles. Some sort of flocculating agent such as an organic polyelectrolyte is usually required to maintain efficient solids capture in the clarifier.

About one pound of dry sludge will be generated per pound of carbon added. If regeneration is practiced, carbon sludge is reactivated and reused with only a small portion removed to prevent buildup of inert material. PAC physicochemical process systems are reasonably reliable. In fact, PAC systems can be used to improve process reliability of existing systems.

Additional information on carbon adsorption and combined biological and physicochemical PACT process systems can be found in Refs. (22–31).

2. CARRIER-ACTIVATED SLUDGE PROCESSES (CAPTOR AND CAST SYSTEMS)

There has been a substantial interest in recent years in the potential benefits of high biomass wastewater treatment. The major obstacle for achieving this has been the inability of biosolids separation in secondary clarifiers. For the most part, this has been overcome by using various forms of support media or carriers that have the ability to attach high concentrations of aerobic bacterial growth (32–34). The increase in immobilized biomass reduces the process dependence on secondary settling basins for clarification. In such hybrid systems where attached growth coexist with suspended growth one gets more stable systems that possess the combined advantages of both fixed and suspended growth reactors.

2.1. Advantages of Biomass Carrier Systems

The performance of carrier systems is dependent on the amount of attached biomass, the characteristics of attached and suspended microorganisms and the type of carriers. The advantages of such hybrid systems are:

(a) Heterogeneity of the microbial population. This is brought about by the differences in the microhabitat of organisms attached to the surface of a carrier and those in the bulk of the solution with respect to pH, ionic strength, and concentration of organics (35–39).
(b) Increased persistence in reactor. This leads to increase in biomass of organisms, reduction of hydraulic retention time and thus smaller reactor volumes (40–42).
(c) Higher growth rate (43–45).

(d) Increased metabolic activity. This leads to increase in respiration and substrate utilization, hence higher removal rates (46–49).
(e) Better resistance to toxicity (50–53).

2.2. The CAPTOR Process

One interesting concept of hybrid systems is the CAPTOR process developed jointly by the University of Manchester Institute of Science and Technology (UMIST) and Simon-Hartley, Ltd., in the United Kingdom. This high biomass approach uses small reticulated polyurethane pads as the bacterial growth medium (54). The pads are added to standard activated sludge aeration reactor, and the system is operated without sludge recycle, essentially combining suspended growth with a fixed film in one process. Excess growth is removed from the pads by periodically passing them through specially designed pressure rollers.

The British Water Research Centre (WRC) and Severn-Trent Water Authority conducted a full-scale evaluation of the CAPTOR process for upgrading the activated sludge plant at the Freehold Sewage Treatment Works, in the West Midlands area of England, to achieve year-round nitrification. This full scale study was jointly sponsored by the US Environmental Protection Agency (55, 56).

2.3. Development of CAPTOR Process

As mentioned earlier, the CAPTOR process originated from research work on pure systems in the Chemical Engineering Department of UMIST. Single strands of stainless steel wire were woven into a knitted formation and then crushed into a sphere of about 6 mm (0.25 in.) diameter. These particles of known surface area were used for modeling liquid-fluidized bed systems. From this work derived the idea of using porous support pads for growing biomass at high concentrations that could be used in wastewater treatment systems. The idea was jointly developed and patented by UMIST and their industrial partner Simon-Hartley, Ltd. The present form of the CAPTOR process uses $25 \times 25 \times 12$ mm ($1 \times 1 \times 0.5$ in.) reticulated polyether foam pads containing pores nominally of about 0.5 to 0.9 mm (0.02 to 0.035 in.) diameter and 94% free space (57–59).

2.4. Pilot-Plant Study

The conducted pilot-plant work indicated that it was possible to achieve the following (55, 56):

(a) Biomass concentrations of 7000 to 10,000 mg/L.
(b) Waste sludge concentrations of 4% to 6% dry solids using a special pad cleaner.
(c) Improved oxygen transfer efficiencies.
(d) High BOD volumetric removal rates.

2.5. Full-Scale Study of CAPTOR and CAST

The full-scale evaluation of the CAPTOR process was undertaken at the Freehold Sewage Treatment Works near Stourbridge, West Midlands. The Freehold plant did not achieve any nitrification in the winter and only partial nitrification in the summer. Freehold's activated

sludge system consisted of five trains equipped with tapered fine bubble dome diffusers arranged in a grid configuration. The system was modified as shown in Fig. 16.2 to split the wastewater flow into two equal volumes. Half went to two trains that were modified by adding CAPTOR pads to the first quarter of two aeration basins, and the other half went to two trains that remained unaltered and served as a control. The CAPTOR modified trains were each equipped with a CAPTOR pad cleaner (Fig. 16.3), and the CAPTOR pads were prevented from escaping into the remainder of the experimental system aeration basins by screens placed at the effluent ends of the CAPTOR zones.

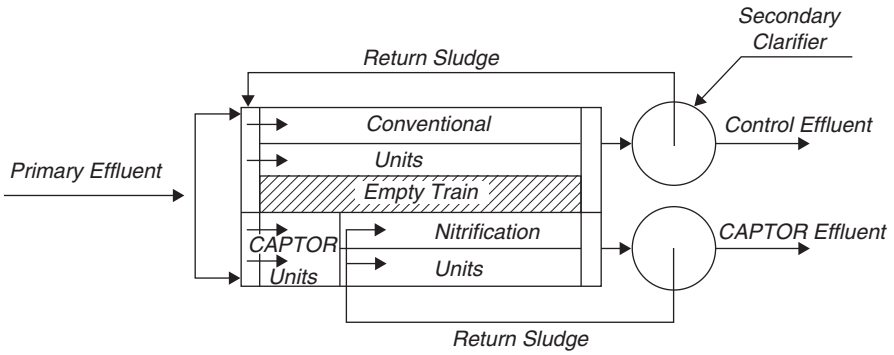


Fig. 16.2. Schematic of treatment plant showing incorporation of CAPTOR (56).

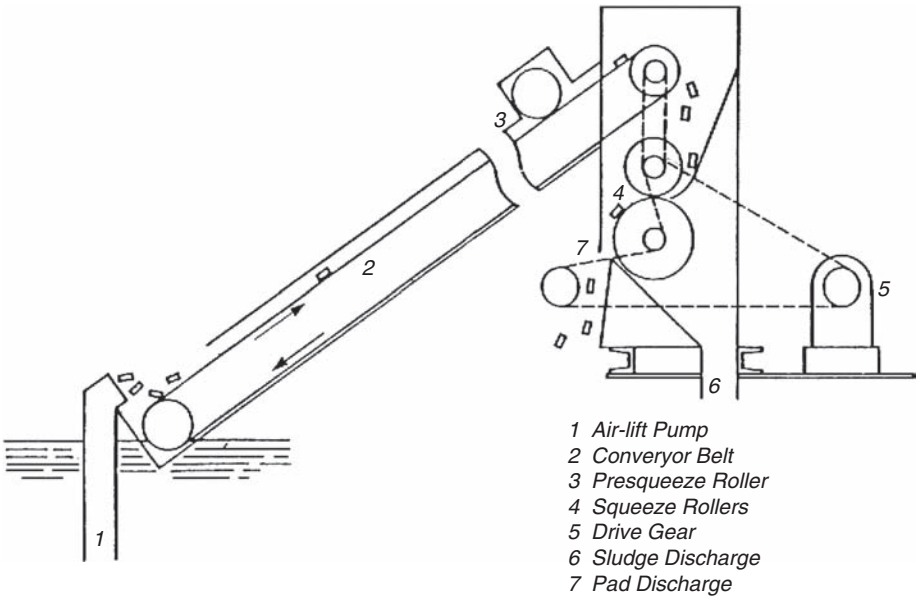


Fig. 16.3. CAPTOR pad cleaner (56).

The Simon-Hartley design predicted that, with a concentration of 40 pads/L, an annual average removal of 75% of the BOD_5 coming into the plant could be achieved in the CAPTOR zones, resulting in a reduced food-to-microorganism (F/M) loading on the follow-on activated sludge stage of 0.08 kg BOD_5/day/kg MLSS. With the reduced load, it was predicted that the modified system would achieve year-round nitrification with an effluent ammonia nitrogen concentration of 5 mg/L or less (56).

2.5.1. Full-Scale Plant Initial Results

The Freehold modified CAPTOR activated sludge system was put in operation and immediately encountered a major problem. The CAPTOR pads floated on the surface of the tanks and would not become incorporated into the tank liquor. A solution was found by removing three of the seven longitudinal rows of fine bubble diffusers in the CAPTOR aeration basins. This was done to create a spiral roll in the tanks, which leads to areas of rising and failing liquid with quite large channels down which the pads can fall. The spiral roll modification provided the necessary falling zone and produced complete mixing of the CAPTOR pads.

Another problem that occurred was mal distribution of the pads. The flow of wastewater tended to push the CAPTOR pads to the outlet of their zones, resulting in a concentration of 50 to 60 pads/L at the outlet and only 10 to 20 pads/L at the inlet end.

One other disturbing feature was the rapid deterioration in the CAPTOR pads. The CAPTOR pads used initially were black and were wearing at such a rate that they would not have lasted for more than 3 years, rendering the process uneconomical.

It had also become evident by this time that with the Freehold wastewater it would be possible to achieve the concentration of 200 mg biomass/pad predicted in the design. However, it was found that if the biomass was allowed to grow beyond 180 mg/pad, the biomass in the center of the pad became anaerobic. The control of pad biomass was difficult because the pad cleaners provided were not reliable and were situated at the CAPTOR zone inlets whereas most of the pads gravitated to the outlet ends of the zones.

During this early period, while the above problems were being tackled on the full-scale plant, there were some occasions when the effluent from the CAPTOR units was reasonable (BOD removals of 40% to 50%), but BOD removal never approached the average of 75% predicted based on the earlier pilot-plant results. Poor BOD removals were being experienced because the suspended solids concentration in the effluent was always high (>80 mg/L).

Consequently more pilot-scale studies were used to find solutions to the operating problems described above before attempting further full-scale evaluation at Freehold.

2.5.2. Pilot-Scale Studies for Project Development

It was decided to evaluate two variations of the CAPTOR process. The new variation differed from the original CAPTOR in that the pads were placed directly into the mixed liquor of the activated sludge aeration tank rather than in a separate stage before the activated sludge tank. WRC named this process variation CAST (CAPTOR in activated sludge treatment). The CAST system had been applied to upgrade several overloaded wastewater treatment plants in Germany and France, and was found to be useful in improving the treatment efficiency and plants performance (60–62).

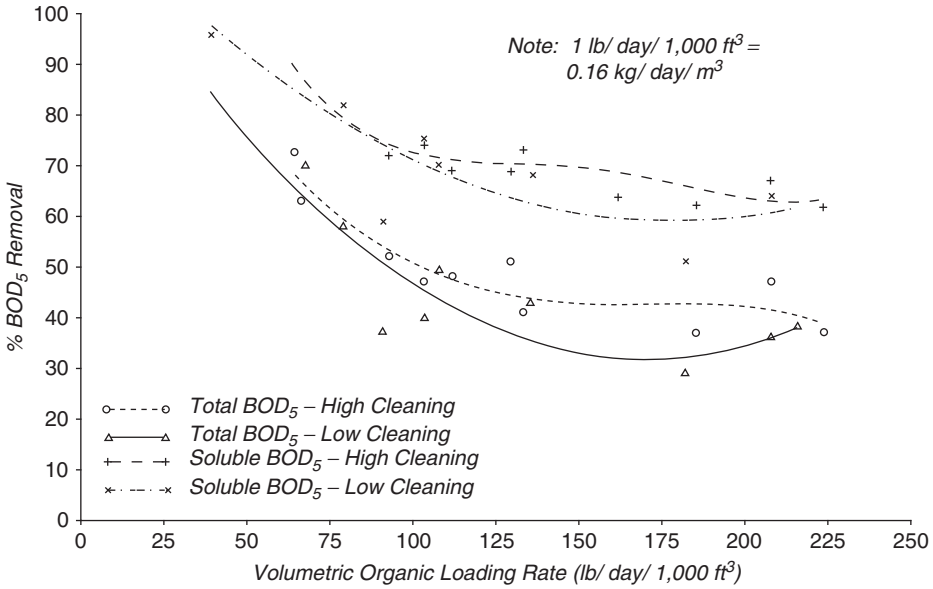


Fig. 16.4. Pilot-scale CAPTOR BOD_5 removals as a function of organic loading rate (56).

In addition, a single aeration tank filled with 40 CAPTOR pads/L, was fed effluent from the above activated sludge control unit to assess the potential of CAPTOR as a second-stage nitrification process. Neither pad cleaning nor final clarification was necessary with this process variation because of the low sludge yields characteristic of nitrifier growth.

Studies were conducted using two well-mixed CAPTOR tanks in series. A range of loading and pad cleaning rates were used to evaluate process removal capabilities for CAPTOR. The intermediate effluent was used as a measure of process efficiency of the primary reactor and the final effluent for the entire system. This permitted plotting (Fig. 16.3) of % BOD_5 removal (total and soluble) vs. volumetric organic loading rate over the range of 1 to 3.5 kg BOD_5/day/m^3 (62 to 218 lb/day/1, 000 ft^3). High and low pad cleaning rates are differentiated in Fig. 16.4 as $\geq$16% and $<$16% of the total pad inventory/day, respectively (56).

Total BOD_5 removal efficiency was less than soluble $B0D_5$ removal efficiency because of the oxygen demand exerted by the biomass solids lost in the process effluent. The higher pad cleaning rates are believed to have contributed to the improved total and soluble BOD removals shown in Fig. 16.4, although low bulk liquid DO's may have adversely affected removals on some of the low cleaning runs. Low cleaning rates ($<$16%/day) were detrimental to soluble $BO0_5$ removal efficiency because of a gradual decline in activity of the biomass remaining in the pad. Cleaning rates greater than 24%/day, however, resulted in reduced biomass levels in the pads and a reduction in performance.

The problem of mal distribution of CAPTOR pads in the aeration tank (i.e., crowding of pads into the effluent end of the tank when operated in plug flow fashion as at Freehold) was solved by modifying the flow pattern to transverse flow (across the width of the tank rather

than down the length). When implemented later at Freehold, this pattern resulted in a fourfold decrease in flow velocity.

Several mixing intensities and diffuser arrangements were tried to decrease biomass shedding into the process effluent. It became obvious; however, that production of effluent biomass solids was not significantly affected by changes in mixing intensity or diffuser arrangement. High effluent suspended solids proved to be far more dependent on pad cleaning rate, biochemical activity of the biomass, and biomass growth directly in the liquor.

Using the transverse flow scheme and a regular pad cleaning regimen, CAPTOR process performance was similar to that experienced in the small tanks. Operating parameters and process performance are summarized in Table 16.2 for two different volumetric loading rates (56).

Respiration studies conducted on pads indicated that biomass held within the pads respires at up to 40% to 50% less than equivalent biomass in free suspension. Any increase in net biomass concentration achieved in a CAPTOR reactor above that in a conventional activated sludge reactor may not produce noticeable benefits, therefore, because of the lower specific activity. These observations suggest that diffusion limitations were occurring in the CAPTOR pads.

The CAST variation of CAPTOR was operated in conjunction with a final clarifier to settle the mixed liquor solids component of the total biomass inventory and return it to the aeration tank. CAPTOR pads and biomass retained therein were kept in the reactor by screens. Operating and performance data are compared in Table 16.3 for the CAST unit and the parallel

Table 16.2
Pilot-scale operating conditions and process performance (56)

	Period			
Parameter	1		2	
Volumetric loading (lb BOD_5/day/1, 000 ft^3)[a]	113		213	
HRT (hr)	2.32		1.52	
Pads/L	40		40	
Biomass/pad (mg)	121		126	
Equivalent MLSS (mg/L)	4.840		5.040	
F/M loading (kg BOD_5/day/kg MLSS)	0.37		0.68	
SRT (days)	3.23		1.72	
DO (mg/L)	4.2		4.7	
	In	Out	In	Out
Total BOD_5 (mg/L)	175	93	216	129
Soluble BOD_5 (mg/L)	86	24	85	33
SS (mg/L)	116	120	178	160
Total BOD_5 removal (%)	47		40	
Soluble BOD_5 removal (%)	72		61	
SS removal (%)	−3		10	

[a] $lb/day/1000\ ft^3 = 0.016\,kg/day/m^3$.

Table 16.3
Pilot-scale CAST and activated sludge operating conditions and performance (56)

Parameter	System			
	CAST		Activated Sludge	
Volumetric loading (lb BOD_5/day/1,000 ft^3)[a]	148		148	
HRT (h)	1.8		1.8	
Pads/L	34		–	
Biomass/pad (mg)	116		–	
Equivalent MLSS in pads (mg/L)	3,930		–	
MLSS in suspension (mg/L)	3,720		6,030	
Total MLSS (mg/L)	7,650		6,030	
F/M loading (kg BOD_5/day/kg total MLSS)	0.31		0.39	
SRT, based on total MLSS (days)	3.6		3.0	
DO (mg/L)	2.5		3.0	
	In	Out	In	Out
Total BOD_5 (mg/L)	178	12	178	20
Soluble BOD_5 (mg/L)	101	5	101	4
SS (mg/L)	121	15	121	23
Total BOD_5 removal (%)	93		89	
Soluble BOD_5 removal (%)	95		96	
SS removal (%)	88		81	

[a] 1 lb/day/1000 ft^3 = 0.016 kg/day/m^3.

activated sludge control unit for a 25-day period when the volumetric loadings and hydraulic residence times (HRT) for both units were identical.

In the nitrification experiments conducted on the CAPTOR process, the biomass concentrations per pad ranged from 99 to 124 mg. This is within the range of 100 to 150 mg/L reported by other researchers (63). With a pad concentration of 40/L, equivalent MLSS levels varied from 3,960 to 4,960 mg/L. Liquor DO concentrations were maintained between 6.4 and 8.4 mg/L, and liquor temperature ranged from 11.50 °C to 6.5 °C.

Secondary effluent from the control activated sludge pilot unit used in the CAST experiments was applied to the nitrification reactor over a range of loading conditions. Essentially complete nitrification was achieved at TKN and ammonia nitrogen loadings of approximately 0.25 kg/day/m^3 (15.6 lb/day/1, 000 ft^3) and 0.20 kg/day/m^3 (12.5 lb/day/1, 000 ft^3), respectively.

2.5.3. Full-Scale Plant Results after Modifications

Following the successful testing of the transverse mixing arrangement in the pilot-scale study, the two Freehold CAPTOR trains were modified. The modifications involved the following (56):

(a) Splitting each of the CAPTOR trains, C1 and C2, into two compartments, C1A and C1B and C2A and C2B, as shown in Fig. 16.5.
(b) Feeding influent flow along long weirs at the side of the trains instead of at the narrow inlet ends.

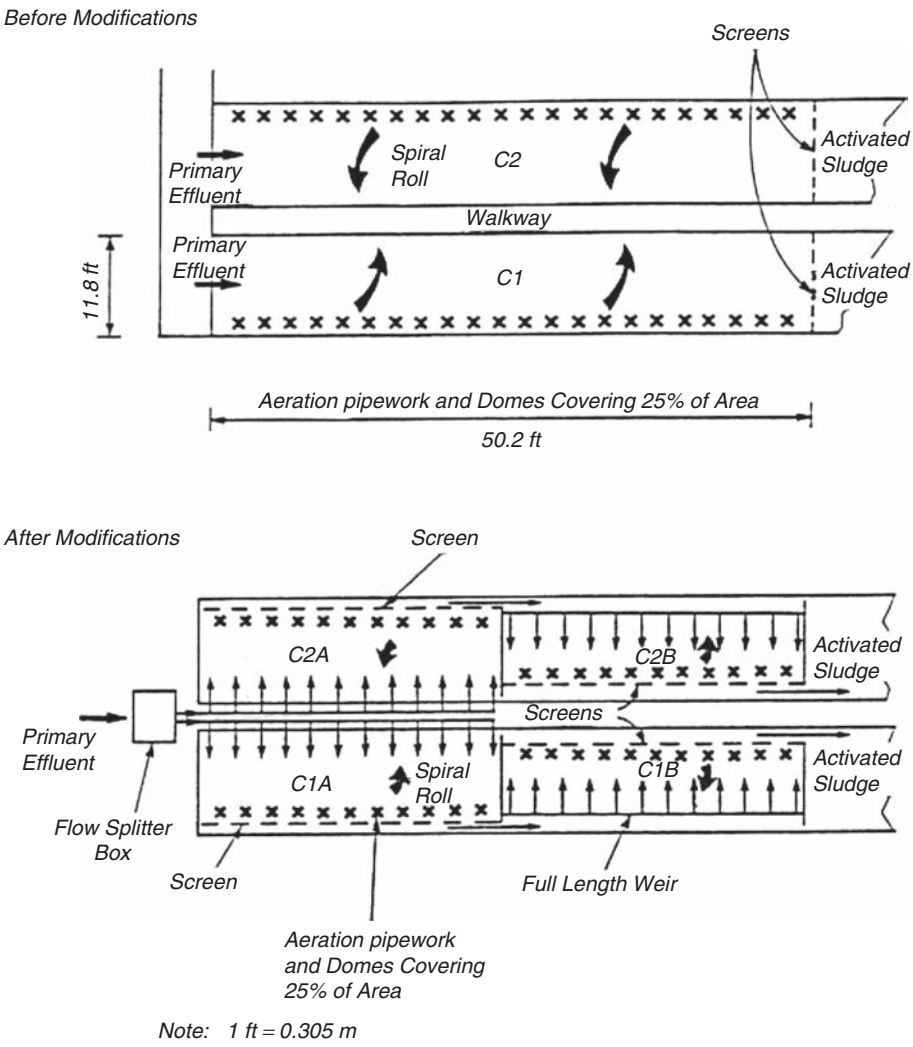


Fig. 16.5. Modifications to full-scale CAPTOR system flow pattern (56).

(c) Modifying the aeration pipe work to place all three rows of dome diffusers directly below the outlet screens (covering about 25% of the width of the tanks), thereby creating a spiral roll of pads and liquid counter-current to the flow of wastewater entering along the weirs on the sidewalls.
(d) Installing two extra pad cleaners so that each CAPTOR sub-unit was provided with a cleaner.
(e) Installing fine screens at the outlet from the primary clarifiers to reduce the quantity of floating plastic material entering the CAPTOR units that created problems with the cleaners.

The objective of the first three modifications was to achieve uniform mixing of the pads in the CAPTOR units and prevent the situation that had occurred previously where high concentrations of pads (50 to 60 pads/L) collected at the outlet end and very low concentrations (10 to

20 pads/L) at the inlet end. Pads were removed from the tanks during the modifications. After the modifications were completed, the number of pads in each compartment was equalized at about 35/L.

The changes were completely successful in obtaining uniform distribution and complete mixing of the CAPTOR pads. A lithium chloride tracer test conducted on the modified tanks indicated that no dead zone was occurring in the "eye" of the roll. Formation of floating pad rafts (which had occurred at the outlet end of the tank with the original arrangement) was completely eliminated. The modifications, however, had no effect on the high level of suspended solids present in the liquor. The modified CAPTOR system was operated at an average volumetric loading rate of 1.24 kg BOD_5/day/m^3 (77 lb/day/1, 000 ft^3), an average HRT (excluding sludge recycle) of 2.55 h and an overall biomass concentration of 4830 mg/L.

The CAST variation of the CAPTOR process, which had exhibited somewhat better performance than conventional activated sludge in the small tank experiments, was also field evaluated at Freehold. The CAPTOR trains were further modified so that return sludge could be introduced to the CAPTOR zones (35 pads/L), providing an activated sludge component throughout the entire aeration tanks, not just in the nitrification stage. The average volumetric organic loadings and HRTs (excluding sludge recycle) were 1.11 kg BOD_5/day/m^3 (69 lb/day/1,000 ft^3) and 3.40 h, respectively.

Performance data summarized in Tables 16.4 and 16.5 indicate that the CAST system exhibits somewhat better performance than the CAPTOR version. In the CAST process the removal of soluble BOD_5 is 96% compared to 90% in CAPTOR; the removal of total BOD_5 is 88% compared to 83%; and the removal of SS is about the same at about 78%.

Table 16.4
Full-scale modified CAPTOR performance results (56)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
Total BOD_5	128	22	83
Soluble BOD_5	40	4	90
SS	138	32	77
NH_4^+-N	24	24.4	0

Table 16.5
Full-scale modified CAST performance results (56)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
Total BOD_5	138	16	88
Soluble BOD_5	56	2	96
SS	120	27	78
NH_4^+-N	26.7	17.2	36

2.5.4. Overall Conclusions

The USEPA conclusions and recommendations for the CAPTOR/CAST treatment systems are as follows (55, 56, 64):

(a) In the initial phase when the CAPTOR process was installed at the Freehold Sewage Treatment Works, several problems were immediately evident. There were major problems with respect to pad mixing, suspension, and distribution and the process performance was adversely affected by the high level of suspended solids in the CAPTOR stage effluent. The problems of pad mixing and distribution were solved by pilot- and full-scale development work.
(b) The performance of the CAPTOR process was still adversely affected by the high level of suspended solids in the CAPTOR stage effluent after correction of the pad mixing, suspension, and distribution problems. This prevented the achievement of nitrification in the follow-on activated sludge stage.
(c) The presence of CAPTOR pads in the tank liquid did not improve oxygen transfer efficiency.
(d) The durability of the CAPTOR pads was solved by switching to different pads.
(e) The peak biomass concentration in the pads is unpredictable. It does not appear to be related to the BOD concentration of the wastewater. There were indications in the various studies, however, that the frequency of pad cleaning (and, hence, the biomass/pad concentration) was critical to the performance of the process. Regular pad cleaning is essential to prevent anaerobic conditions from developing in the pads.
(f) It is possible to raise the biomass concentration in a CAPTOR stage to 6,000 to 8,000 mg/L, but the respiration rate of the biomass in the pads is lower than the respiration of the same biomass if freely suspended and less than that of normal activated sludge. These data suggest that the geometry of the CAPTOR pads results in diffusion limitations, which demands further pad design improvement to enhance the potential for economic use of the CAPTOR process in wastewater treatment.
(g) The CAST variation of the CAPTOR process performs well.
(h) CAPTOR has the potential as an add-on package for tertiary nitrification.
(i) The CAPTOR option was projected to be more cost effective than extending the activated sludge plant for upgrading Freehold to complete year-round nitrification.
(j) For CAPTOR and CAST to achieve their full potential, as predicted by the pilot-scale studies, further design development and improvements are needed.

3. ACTIVATED BIO-FILTER (ABF)

3.1. Description

Activated bio-filters (ABF) are a recent innovation in the biological treatment field. This process consists of the series combination of an aerobic tower (bio-cell) with wood or other packing material, followed by an activated sludge aeration tank and secondary clarifier. Settled sludge from the clarifier is recycled to the top of the tower. In addition, the mixture of wastewater and recycle sludge passing through the tower is also recycled around the tower, in a similar manner to a high rate trickling filter. No intermediate clarifier is utilized. Forward flow passes directly from the tower discharge to the aeration tank (Fig. 16.6). The use of the two forms of biological treatment combines the effects of both fixed and suspended growth processes in one system. The microorganisms formed in the fixed growth phase are passed along to the suspended growth unit, whereas the suspended growth microorganisms

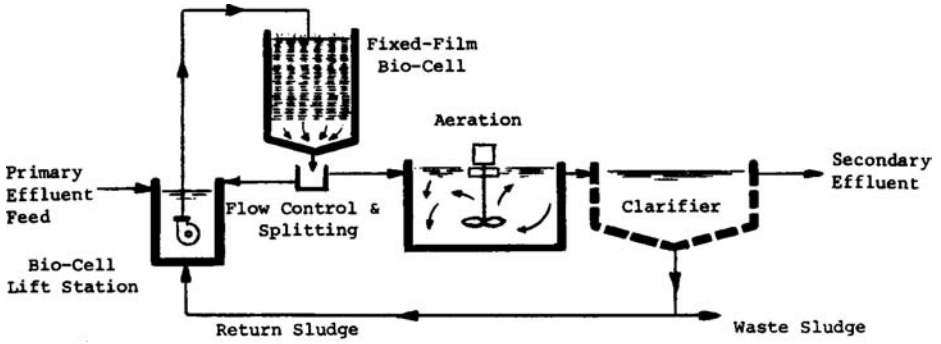


Fig. 16.6. ABF process flow diagram (65).

are recycled to the top of the fixed media unit (65). This combination of the two processes results in the formation of a highly stable system that has excellent performance and good settling biological floc when treating wastewaters that have variable loads (66).

The bio-media in the bio-cell consists of individual racks made of wooden laths fixed to supporting rails. The wooden laths are placed in the horizontal direction, permitting wastewater to pass downward, and air horizontally and vertically. The horizontal surfaces reduce premature sloughing of biota. Droplet formation and breakup induced by wastewater dripping from lath to lath enhances oxygen transfer. Other types of material for the bio-media have also been reported by other researchers and equipment manufacturers (67–70). The aeration basin is a short detention unit that can be designed for either plug flow or complete mix operation. The effluent from the aeration basin passes to a secondary clarifier where the activated sludge is collected and recycled to the top of the bio-cell tower and to waste.

ABF units can be used for the removal of either carbonaceous material or for carbonaceous removal plus nitrification by appropriately modifying the detention time of the aeration basin. When nitrification is desired, the bio-cell acts as a first-stage roughing unit and the aeration basin as a second-stage nitrification unit (71, 72). ABF bio-cells can be either rectangular or round. Various types of aeration equipment can be used in the aeration system, including both surface and diffused aerators. The detention time of the aeration tank can be modified, depending on influent quality and desired effluent quality. ABF units can be supplied with mixed media effluent filters for enhanced treatment.

3.2. Applications

Activated bio filters can be used for treating municipal wastewater and biodegradable industrial wastewater. ABF systems are especially useful where (65, 66):

(a) Both BOD_5 removal and nitrification are required.
(b) Land availability is low.
(c) Raw wastewater organic loadings fluctuate greatly, because of its ability to handle shock conditions.
(d) Existing trickling filter facilities and overloaded existing secondary plants need to be upgraded at reduced cost.

Table 16.6
Performance of BAF systems (65)

Parameter	Influent, mg/L	Effluent, mg/L	Removal, %
BOD_5	153	14	91
COD	330	58	82
TSS	222	20	91
NH_4^+-N*	20	1	90

* When used for nitrification.

A typical ABF application is the Burwood Beach Wastewater Treatment Works in Australia (73). The plant was upgraded in the 1990s using ABF at a cost of $48 million. The facility currently serves a population of 180,000 with a flow of 43 ML/day and has the capacity to treat 53 ML/day, for a population of 220,000 in the year 2020. The Biofilter is 30 m in diameter and has a design organic loading of 3.2 kg BOD_5/m^3/day. The aeration tank is designed for 1.5 hours of hydraulic detention time. The plant has been in operation for around 10 years producing an effluent that is consistently within the required EPA set limits.

3.3. Design Criteria

The design criteria for the ABF system are reported to be as follows (65, 74, 75):

(a) Bio-cell organic load: 100 to 200 lb BOD_5/day/l000 ft^3
(b) Return sludge rate: 25% to 100%
(c) Bio-cell recycle rate: 0% to 100%
(d) Bio-cell hydraulic load: 1 to 5.5 gpm/ft^2
(e) Aeration basin detention time: 0.5 to 3.0 hours for BOO_5 removal only
    5.8 to 7.5 hours for two-stage nitrification
(f) System F/M: 0.25 to 1.5 lb BOD_5/day/lb MLVSS for BOD removal
    0.18 lb BOD_5/day/lb MLVSS for two-stage nitrification

3.4. Performance

ABF systems are quite stable and highly reliable. They can treat standard municipal, combined municipal/industrial, or industrial wastewaters to BOD_5 and suspended solids levels of 20 mg/L or less. Test study on a package system showed at least 90% removal of BOD_5, TSS and $NH_4^+ - N$ (65). The detailed results are shown in Table 16.6.

Sludge production was reported at 0.25 to 1.0 lb of waste VSS per lb of BOD_5 removed. The mean yield over the course of the study was 0.60 lb VSS per lb of BOD removed.

4. VERTICAL LOOP REACTOR (VLR)

4.1. Description

A vertical loop reactor (VLR) is an activated sludge biological treatment process similar to an oxidation ditch (76, 77). The wastewater in an oxidation ditch circulates in a horizontal loop; the water in a VLR circulates in a vertical loop around a horizontal baffle, as shown in

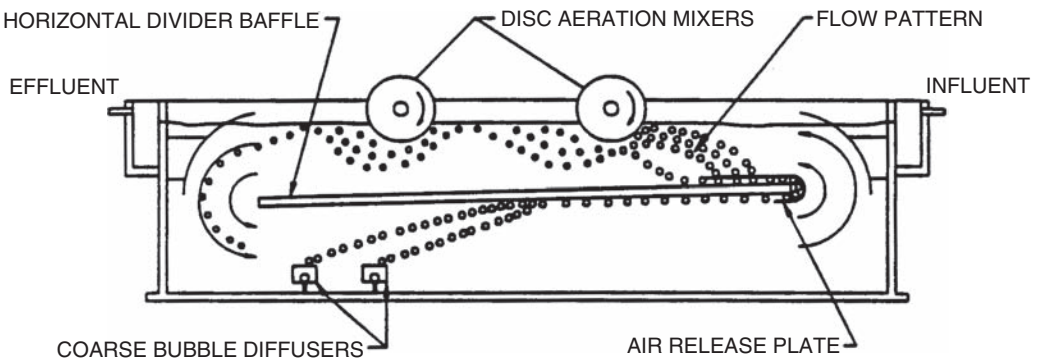


Fig. 16.7. Diagram of the vertical loop reactor (77, 78).

Fig. 16.7 (78). A typical VLR consists of an 18 ft deep concrete or steel basin with a horizontal baffle extending the entire width of the reactor and most of its length. Operating basins are reported to have side-wall depths that range from approximately 10 to 22 ft (79). The length and width of the VLR are determined by the required capacity but, as a rule, the length is at least twice the width. The baffle is generally five to eleven feet below the surface of the water. Because a VLR is typically deeper than an oxidation ditch, the VLR requires less land area.

Aeration in a VLR is provided by coarse bubble diffusers, which are located below the horizontal baffle and by disc aeration mixers. The disc mixers also circulate the wastewater around the baffle at a velocity of 1 to 1.5 ft/s (80). Because the diffusers are positioned below the baffle, the air bubble residence time in a VLR is as much as six times longer than the bubble residence time in a conventional aeration system. This extended bubble contact time increases the process aeration efficiency. Denitrification in an anoxic zone also reduces oxygen requirements.

The VLR process is usually preceded by preliminary treatment such as screening, comminution or grit removal. Secondary settling of the VLR effluent is typically provided by a separate clarifier. An intra-channel clarifier may be used for secondary settling in place of a separate clarifier.

Vertical loop reactors may be operated in parallel or series. When a series of VLRs are used, the dissolved oxygen profile can be controlled to provide nitrification, denitrification and biological phosphorus removal at hydraulic detention times of 10 to 15 h.

4.2. Applications

VLR technology is applicable in any situation where conventional or extended aeration activated sludge treatment is appropriate. The technology is applicable for nitrification and denitrification. Biological phosphorus removal may be incorporated in the system design. Power costs may be lower for a VLR system than for other aerated biological treatment systems, because of improved oxygen transfer efficiency. There are currently more than ten municipal wastewater treatment facilities in the United States with VLRs. One such example

is the City of Willard, OH waste water treatment plant (81). The facility is designed for an average daily flow of 4.5 MGD and is capable of handling a peak flow of 7.2 MGD.

The following advantages have been reported for VLR systems (82):

(a) Land area required for VLRs is about 40% less than for oxidation ditches.
(b) The VLR aeration basin cost is about 30% less than for oxidation ditches.
(c) The multiple tank basin series arrangement is an advantage for facilities with highly variable flow.
(d) VLRs are useful for retrofitting existing basins for plant upgrade to suit increased flows or more stringent effluent requirements.

4.3. *Design Criteria*

The design criteria for the VLR process are reported to be as follows (76):

BOD_5 loading	14 to 22 lb $BOD_5/1{,}000\,ft^3/day$
SRT	17 to 36 day
Detention Time	12 to 24 hours

4.4. *Performance*

The average effluent BOD_5 and TSS concentrations for the five studied operating VLR facilities are 4.2 and 7.1 mg/L, respectively. The average effluent ammonia concentration is 0.8 mg/L. Only one of the VLRs studied was designed for biological phosphorus removal; the average effluent phosphorus concentration for this plant was 1.45 mg/L and alum was added in the final clarifiers. A second VLR facility was not designed for biological phosphorus removal but was required to monitor phosphorus. This plant had an average effluent phosphorus concentration of 2.19 without any chemical addition.

The VLR system is quite reliable. Table 16.7 indicates the percent of time the monthly average effluent concentration of the given pollutants was less than the concentration given in the first column. No significant difference in results was observed between winter and summer data.

Table 16.7
Reliability of the VLR treatment process (76)

Concentration, mg/L	BOD_5*	NH_4^+-N*	TSS*	P*
0.2	0	30	0	2
0.5	0	63	1	10
1.0	0	83	1	24
2.0	20	88	5	63
3.0	71	95	43	93
10.0	97	96	75	100
20.0	100	100	96	100
Number of plants	5	5	5	1

* Percentage of time the monthly average concentration of the pollutant was less than the stated value in the first column.

4.5. EPA Evaluation of VLR

The following summarizes the major findings and conclusions of EPA evaluation of VLRs (77). The information is based on analysis of available information from site visits, a detailed design of a full scale VLR system and information from consultants and manufacturers.

(a) The VLR is a modification of the conventional activated sludge process. The unique features of the process are circulating mixed liquor around a horizontal baffle with a dual aeration system, bubble diffused air beneath the horizontal baffle and disc aerators at the surface of the aeration tank. The process operates as a plug flow reactor with capability for varying dissolved oxygen profiles to achieve biological, phosphorus and nitrogen removal. The VLR process also features a stormwater by-pass design for treatment of high peak to average flows.

(b) There are currently over ten operating VLRs in the US ranging in size from 0.22 to 5.0 MGD.

(c) Performance data from operating VLRs show that this process is capable of achieving effluent carbonaceous biochemical oxygen demand levels of less than 10 mg/L; effluent total suspended solids levels of less than 10 mg/L; and effluent ammonia-nitrogen levels of less than 1.0 mg/L. The process is further capable of achieving total nitrogen and phosphorus removals of 60% to 80%.

(d) The VLR process is applicable for flows ranging from 0.05 to over 10 MGD.

(e) The claimed advantages of this process by the manufacturer include the following:
- Higher dissolved oxygen transfer than conventional equivalent technology.
- Improved response to peak flows because of a stormwater bypass feature.
- A credit for oxygen release because of denitrification with the credit based on 80% denitrification.
- Increased mixed liquor settleability and process stability.

(f) The design criteria for the existing VLRs are conservative. HRTs range from 11.9 to 24 hours. Volumetric loading ranged from 13.6 to 23.1 lb CBOD/1, 000 ft^3. This loading is similar to that used for extended aeration systems and is about 1/3 to 1/2 of that normally used for conventional activated sludge designs.

(g) The VLR technology has been designated as Innovative Technology by the EPA for three plants because of a 20% claimed energy savings.

(h) Based on this assessment, the 20% energy savings over competing technology could not be verified.

(i) The VLR was compared to oxidation ditches as "Equivalent Technology." The results of this comparison indicated:
- The VLR technology produces comparable to slightly improved effluent levels of BOD, TSS and NH3-N than oxidation ditch plants.
- Total removal of phosphorus and total nitrogen are equivalent to oxidation ditches designed for the same level of treatment.
- The energy requirements for aeration were found to be similar to 10% less than for oxidation ditches.
- The land area required for VLRs was found to be approximately 40% less than for oxidation ditches based on equivalent aeration tank loadings.
- The VLR aeration basin cost was found to be approximately 30% less than for oxidation ditches for situations where rock excavation is not required for the deeper VLR basin.

- A definitive comparison of total VLR plant costs to total oxidation plant costs could not be made. Data submitted from both manufacturers's indicated a comparable cost for plants in the 0 to 2 MGD range. The reported VLR costs at plants ranging from 2 to 10 MGD were significantly less than oxidation ditch plant costs. This would be expected because of the modular design and common wall construction of the VLR compared to oxidation ditches.
- The total operation and maintenance costs of the two technologies were found to be similar.

4.6. Energy Requirements

The VLR energy requirements are shown in Fig. 16.8. The requirements are based on the following assumptions (76):

(a) Water Quality
  BOD_5: Influent = 200 mg/L, Effluent = 20 mg/L
  TKN; Influent = 35 mg/L, Effluent = 1 mg/L
(b) Design Basis
  Oxygen transfer efficiency: 2.5 lb O_2/Hp hour
  Nitrification occurs
(c) Operating Parameters
  Oxygen Requirement: 1.5 lb O_2/lb BOD_5 removed
  4.57 lb O_2/lb TKN removed
(d) Type of energy: Electrical

4.7. Costs

Construction costs (1991 dollars, Utilities Index = 392.35) for VLR are shown in Fig. 16.8. To obtain the values in terms of the present 2007 US dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of 539.74/392.35 = 1.38 (83). The operation costs are similar to oxidation ditch type treatment plant.

5. PHOSTRIP PROCESS

5.1. Description

"PhoStrip" is a combined biological-chemical precipitation process based on the use of activated sludge microorganisms to transfer phosphorus from incoming wastewater to a small concentrated substream for precipitation. As illustrated in Fig. 16.9, the activated sludge is subjected to anoxic conditions to induce phosphorus release into the sub-stream and to provide phosphorus uptake capacity when the sludge is returned to the aeration tank. Settled wastewater is mixed with return activated sludge in the aeration tank. Under aeration, sludge microorganisms can be induced to take up dissolved phosphorus in excess of the amount required for growth. The mixed liquor then flows to the secondary clarifier where liquid effluent, now largely free of phosphorus, is separated from the sludge and discharged. A portion of the phosphorus-rich sludge is transferred from the bottom of the clarifier to a thickener-type holding tank: the phosphate stripper. The settling sludge quickly becomes anoxic and, thereupon, the organisms surrender phosphorus, which is mixed into the supernatant. The phosphorus-rich supernatant, a low volume, high concentration substream, is removed from

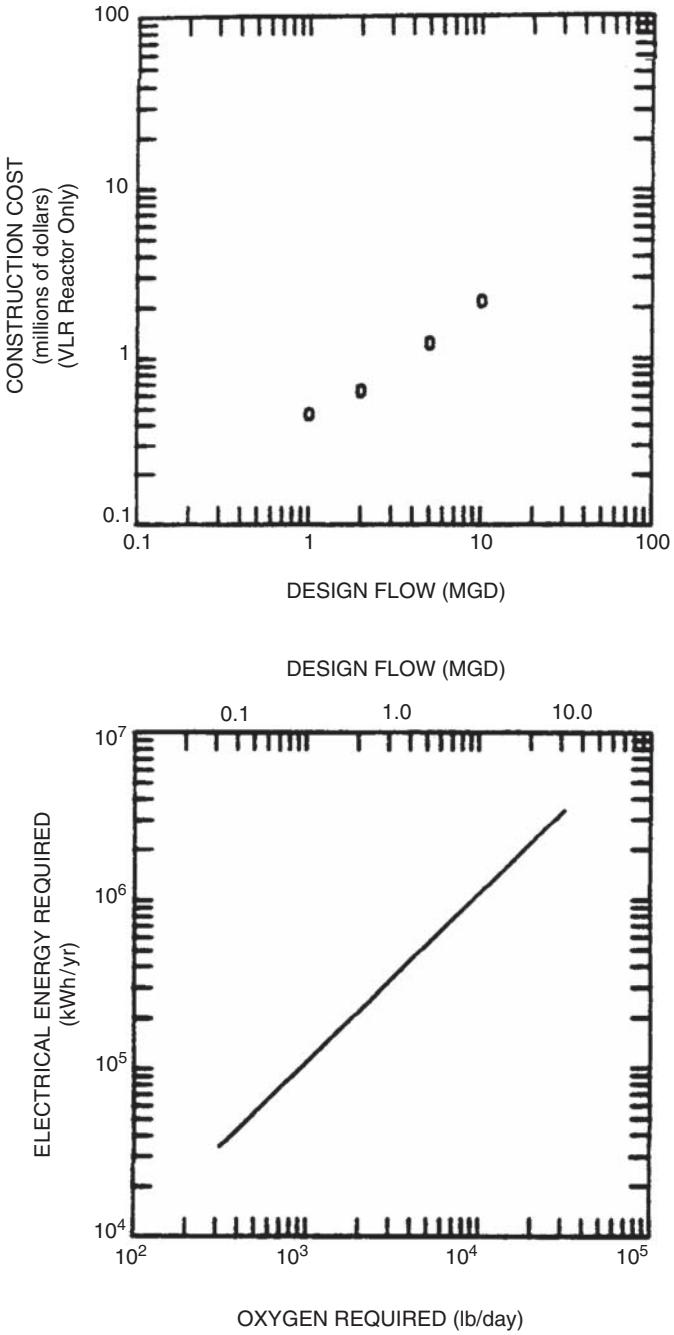


Fig. 16.8. VLR energy requirements and construction ost (76, 77).

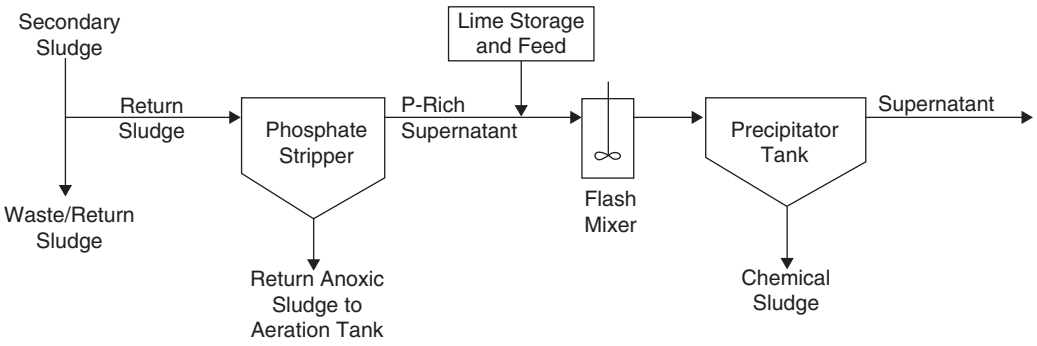


Fig. 16.9. PhoStrip process flow diagram (65).

the stripper and treated with lime for phosphorus precipitation. The thickened sludge, now depleted in phosphorus, is returned to the aeration tank for a new cycle (65).

The PhoStrip process has demonstrated a compatibility with the conventional activated sludge process and is compatible with its modifications. The process can operate in various flow schemes, including full or split flow of return activated sludge through the phosphate stripper, use of an elutriate to aid in the release of phosphorus from the anoxic zone of the stripper, or returning lime-treated stripper supernatant to the primary clarifier for removal of chemical sludge.

This technique is a new development in municipal wastewater treatment and has been demonstrated in pilot plant and full-scale studies. Notable large scale evaluations have been conducted at Seneca Falls, New York and, more recently, Reno/Sparks, Nevada. Nearly a dozen commercial installations are reported to be in the operational phase.

5.2. Applications

This method, which involves a modification of the activated sludge process, can be used in removing phosphorus from municipal wastewaters to comply with most effluent standards. Direct chemical treatment is simple and reliable, but it has the two disadvantages of significant sludge production and high operating costs. The PhoStrip system reduces the volume of the substream to be treated, thereby reducing the chemical dosage required, the amount of chemical sludge produced, and associated costs. Lime is used to remove phosphorus from the stripper supernatant at lower pH levels (8.5 to 9.0) than normally required. The cycling of sludge through an anoxic phase may also assist in the control of bulking by the destruction of filamentous organisms to which bulking is generally attributed (65).

On the negative side, it should be pointed out that more equipment and automation, along with a greater capital investment, are normally required than for conventional chemical addition systems. Because this method relies on activated sludge microorganisms for phosphorus removal, any biological upset that hinders uptake ability will also affect effluent concentrations. It has been found that sludge in the stripper tank is very sensitive to the presence of oxygen. Anoxic conditions must be maintained for phosphorus release to occur.

Table 16.8
Typical design criteria for the PhoStrip process (74)

Design parameter	Unit	Value
Food-to-microorganisms ratio (F/M)	lb BOD_5/lb MLSS/day	0.1–0.5
Solids retention time (SRT)	Day	10–30
Mixed liquor suspended solids (MLSS)	Mg/L	600–5000
Hydraulic retention time in stripper (t)	Hour	8–12
Hydraulic retention time in aeration tank (t)	Hour	4–10
Return activated sludge (RAS)	% of influent	20–50
Internal recycle (stripper underflow)	% of influent	10–20

5.3. Design Criteria

The fraction of the total sludge flow that must be processed through the stripper tank is determined by the phosphorus concentration in the influent wastewater to the treatment plant and the level required in the treated effluent. Required detention time in the stripper tank ranges from five to fifteen hours. Typical phosphorus concentrations produced in the stripper are in the range of 40 to 70 mg/L. The volume of the phosphorus-rich supernatant stream to be lime treated is 10% to 20% of the total flow (65). Typical design criteria for the PhoStrip process are shown in Table 16.8 (74).

5.4. Performance

Pilot and full-scale studies of the process have shown it to be capable of reducing the total phosphorus concentration of typical municipal wastewaters to 1.5 mg/L (74) or even to 0.5 mg/L or less (75). A plant-scale evaluation of the method treating 6 MGD of municipal wastewater at the Reno/Sparks Joint Water Pollution Control Plant in Nevada demonstrated satisfactory performance for achieving greater than 90% phosphorus removal. Results showed that the process enhanced the overall operation and performance of the activated sludge process, because it produced a more stable, better settling sludge. Regular maintenance of mechanical equipment, including pumps and mixers, is necessary to ensure proper functioning of entire system.

5.5. Cost

5.5.1. Construction Cost

Construction costs (1980 dollars, Utilities Index = 277.60) for PhoStrip are shown in Fig. 16.10. To obtain the values in terms of the present 2004 US dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of 506.13/277.60 = 1.82 (83). Construction costs include: stripper (10 h detention time at 50% of return sludge); flash mixer; flocculator/clarifier; thickeners; lime feed and storage facilities (65).

5.5.2. Operation and Maintenance Cost

The electrical energy required for operation of pumps, lime mixing equipment and clarifiers is shown in Fig. 16.11. Operation and maintenance costs (1980 dollars, Utilities

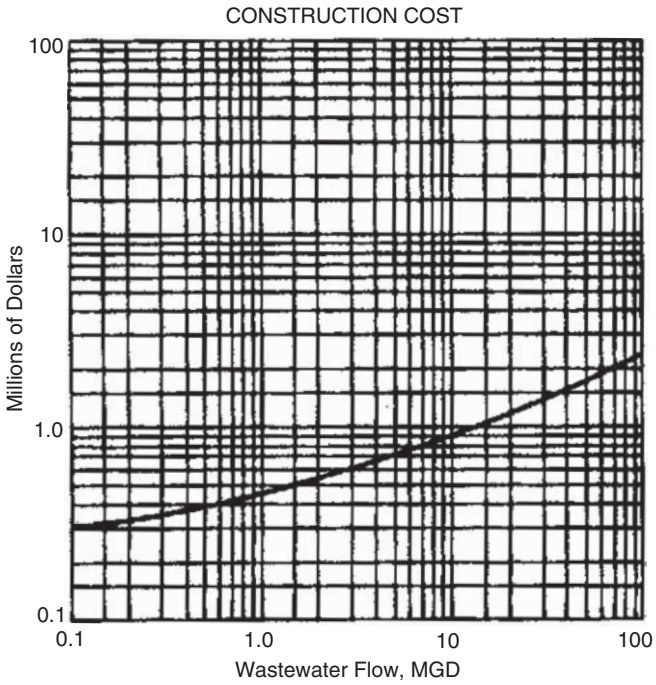


Fig. 16.10. PhoStrip construction cost (65).

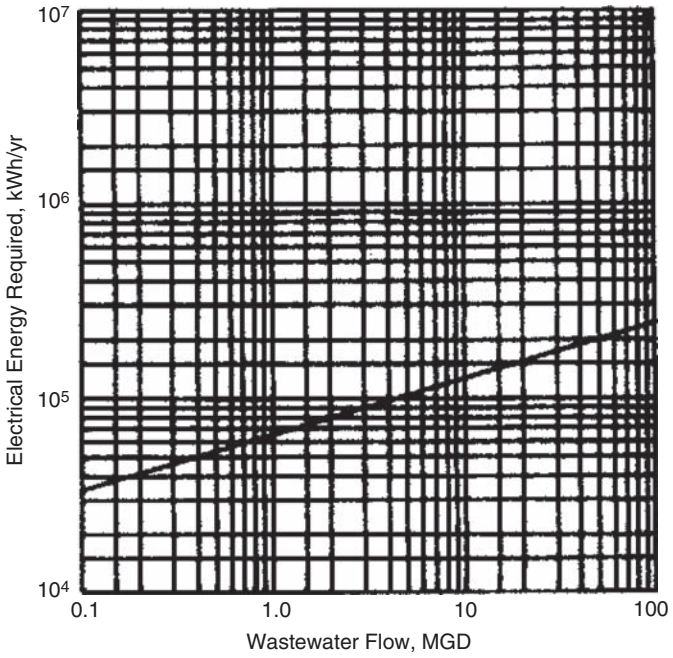


Fig. 16.11. PhoStrip electrical energy requirement (65).

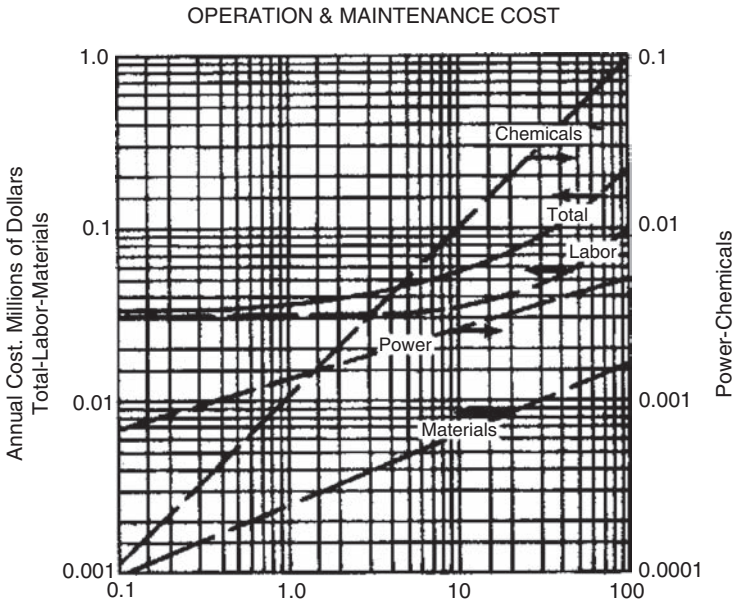


Fig. 16.12. PhoStrip operation and maintenance cost (65).

Index = 277.60) for PhoStrip are shown in Fig. 16.12. To obtain the values in terms of the present 2007 US dollars, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of 539.74/277.60 = 1.94 (83). Operation and maintenance costs include: labor for operation, preventive maintenance, and minor repairs; materials to include replacement parts and major repair work; lime and power cost based on the electrical energy requirement shown in Fig. 16.11 (65).

NOMENCLATURE

BOD_5 = 5-day biochemical oxygen demand, mg/L
COD = chemical oxygen demand, mg/L
DO = dissolved oxygen, mg/L
F/M = food to microorganism loading, kg BOD_5/day/kg MLSS
NH_4^+-N = ammonia nitrogen, mg/L
MLSS = mixed liquor suspended solids, mg/L
MLVSS = mixed liquor volatile suspended solids, mg/L
PACE = effluent PAC concentration, mg/L
PACI = inffluent PAC concentration, mg/L
PACR = mixed liquor PAC concentration in the reactor, mg/L
HRT = hydraulic retention time, day (or hr)
SRT = design solids retention time, day
SS = suspended solids, mg/L
TSS = total suspended solids, mg/L

REFERENCES

1. L. K. Wang, New dawn in development of adsorption technologies, *The 20th Annual Meeting of the Fine Particle Society Symposium on Activated Carbon Technology*, Boston, MA, Aug. (1989).
2. L. K. Wang and Y. Li, Sequencing batch reactors. In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ (2009).
3. L. K. Wang and L. Kurylko, *Sequencing Batch Liquid Treatment*. U. S. Patent and Trademark Office, Washington, DC. U. S. Patent #. 5,354,458, Oct. (1994).
4. L. K. Wang, P. Wang, and N. L. Clesceri, Groundwater decontamination using sequencing batch processes, *Water Treatment*, 10, 2, 121–134 (1995).
5. M. Krofta, L. K. Wang, and M. Boutroy, *Development of a New Treatment System Consisting of Adsorption Flotation and Filtration*, US Dept. of Commerce, National Technical Information Service, Springfield, VA, USA. Report #.PBS5-20940l/AS, p. 28 (1984).
6. L. K. Wang and R. Menon, Membrane bioreactor. In: *Advanced Biological Treatment Processes*, L. K. Wang, N. K. Shammas, and Y. T. Hung (eds.), Humana Press, Totowa, NJ (2009).
7. L. K. Wang, Removal of organic pollutants by adsorptive bubble separation processes, *1974 Earth Environment and Resources Conference Digest of Technical Papers*, 1, 74, 56–57, Sept. (1974).
8. WEF and ASCE, *Design of Municipal Wastewater Treatment Plants*, WEF Manual of Practice No 8 and ASCE Manual and Report on Engineering Practice No. 76. WEF, Alexandria, VA (1992).
9. T. L. Randall, W. M. Copa, and M. J. Dietrich, Leachate treatment by a powdered activated carbon process, Presented at the *59th Annual Conference of the Water Pollution Control Federation*, Los Angeles, CA, Oct. (1986).
10. K. J. Deeny, J. A. Heidman, and A. J. Condren, *Performance of Activated Sludge Powdered Activated Carbon/Wet Air Regeneration Systems*, EPA 600/S2-90/012, Cincinnati, OH (1990).
11. K. Depuydt and R. Amundson, Solving an ash buildup challenge. Pollution Engineering, p. 73 Dec. (1991).
12. J. A. Meidl, Personal Communication from Zimpro Passavant Environmental Systems, Inc., to P. M. Sutton Oct. (1991).
13. L. K. Wang and Z. Wu, Activated sludge processes. In: *Biological Treatment Processes*, L. K. Wang, N. C. Pereira, Y. T. Hung, and N. K. Shammas (eds.), Humana Press, Totowa, NJ (2009).
14. US EPA, *Nitrogen Control*, Tech. Report # EPA/625/R-93/010, U. S. Environmental Protection Agency, Washington, DC, (1993).
15. L. Beftens, Powdered activated carbon in an activated sludge unit, *Journal of Effluent and Water Treatment*, 9, 129 (1979).
16. N. A. Leipzig, Effectiveness of the powdered activated carbon activated sludge system in removing ammonia from an organic chemical production wastewater, *Proceedings of the 35th Industrial Waste Conference*, pp. 889–897, Purdue University, Lafayette, IN. Ann Arbor, MI, (1980).
17. A. S. Ng and M. K. Stenstrom, Nitrification in powdered-activated carbon- activated sludge process, *Journal of Environmental Engineering*, 113, 1285 (1987).
18. N. K. Shammas, Interaction of temperature, pH, and biomass on the nitrification process, *J. Water Pollution Control Federation*, 58, 1, 52–59, Jan. (1986).
19. L. K. Wang, *Manufacturers and Distributors of Activated Carbons and Adsorption Filters*, p. 33, Zorex Corporation, Pittsfield, MA, Technical Report # P917-5-89-7, May (1989).
20. Editor, Water & wastewater products: 2004 buyer's guide, *Environmental Protection*, p. 163 Dec. (2003).
21. Editor, Water & wastes digest 2004 reference guide, *Water & Wastes Digest*, p. 95, Bolingbrook, IL, Dec. (2003).

22. L. K. Wang, The adsorption of dissolved organics from industrial effluents onto activated carbon, *Journal of Applied Chemistry and Biotechnology*, 25, 7, 491–503 (1975).
23. L. K. Wang, Adsorption, coagulation and filtration make a useful treatment combination, Part I, *Water and Sewage Works*, 123, 12, 42–47, Dec. (1976).
24. L. K. Wang, Adsorption, coagulation and filtration make a useful treatment combination, Part II, *Water and Sewage Works*, 124, 1, 32–36, Jan. (1977).
25. L. K. Wang, *Treatment of Potable Water from Seoul, Korea by Flotation, Filtration and Adsorption*, p. 21, US Dept. of Commerce, National Technical Information Service, Springfield, VA. PB88-200530/AS, March (1988).
26. L. K. Wang, M. H. S. Wang, and J. Wang, *Design, Operation and Maintenance of the Nation's Largest Physicochemical Waste Treatment Plant*, Vol. 1, p. 183, Lenox Institute of Water Technology, Lenox, MA, Report # LIR/03-87-248, March (1987).
27. L. K. Wang, M. H. S. Wang, and J. Wang, *Design, Operation and Maintenance of the Nation's Largest Physicochemical Waste Treatment Plant*, Vol. 2, p. 161, Lenox Institute of Water Technology, Lenox, MA, Report # LIR/03-87/249, March (1987).
28. L. K. Wang, M. H. S. Wang, and J. Wang, *Design Operation and Maintenance of the Nation's Largest Physicochemical Waste Treatment Plant*, Vol. 3, p. 227, Lenox Institute of Water Technology, Lenox, MA, Report # LIR/03-87/250, March.
29. L. K. Wang, *Removal of Heavy Metals, Chlorine and Synthetic Organic Chemicals by Adsorption*, p. 47, Zorex Corporation, Pittsfield, MA, Tech. Report # P917-5-89-8, May (1989).
30. L. K. Wang, Reduction of color, odor, humic acid and toxic substances by adsorption, flotation and filtration, Annual Meeting of American Institute of Chemical Engineers, *Symposium on Design of Adsorption Systems for Pollution Control*, p. 18, Philadelphia, PA, Aug. (1989).
31. L. K. Wang, *The State-of-the-Art Technologies for Water Treatment and Management*, p. 145, United Nations Industrial Development Organization (UNIDO), Vienna, Austria, UNIDO Training Manual # 8-8-95, Aug. (1995).
32. C. Webb, G. M. Black, and B. Atkinson (eds.), *Process Engineering Aspects of Immobilized Cell Systems*, Pergamon Press, Elmsford, NY (1986).
33. J. Tampion and M. D. Tampion, *Immobilized Cells: Principles and Applications*, Cambridge University Press, Cambridge, MA (1987).
34. M. Moo-Young (ed.), *Bioreactor Immobilized Enzymes and Cells-Fundamentals and Applications*, Elsevier Applied Science, New York (1988).
35. C. E. ZoBell, The effect of solid surfaces upon bacterial activity, *Journal of Bacteriology*, 46, 39 (1943).
36. K. L. Sublette, E. H. Snider, and N. D. Sylvester, A review of the mechanism of powdered activated carbon enhancement of activated sludge treatment, *Water Research*, 16, 1075 (1982).
37. R. Z. Maigetter and R. M. Plister, A mixed bacterial population in a continuous culture with and without kaolinite, *Canadian Journal of Microbiology*, 21, 173 (1975).
38. D. Oakley, The retention of biomass in fast flowing systems. In: *Process Engineering Aspects of Immobilised Cell Systems*, C. Webb, O. M. Black, and B. Atkinson (eds.), Pergamon, Elmsford, NY (1986).
39. J. N. Wardell, C. M Brown, D. C. Ellwood, and A. E. Williams, Bacterial growth on inert surfaces. In: *Continuous Culture 8: Biotechnology, Medicine and the Environment*, A. C. R. Dean, D. C. Ellwood, and C. G. T. Evans, Ellis Horwood, Chichester, England (1984).
40. W. J. Jewell, Anaerobic attached film expanded bed fundamentals. In: *Fixed Film Biological Process for Wastewater Treatment*, Y. C. Wu and E. D. Smith (eds.), Noyes Publishing, Park Ridge, NJ (1983).

41. R. J. Shimp and F. K. Pfaender, Effects of surface area and flow rate on marine bacterial growth in activated carbon columns, *Applied Environmental Microbiology*, 44, 471 (1982).
42. W. J. Weber, Jr., M. Pirbazari, and G. L. Melson, Biological growth on activated carbon: an investigation by scanning electron microscopy, *Environmental Science and Technology*, 12, 817 (1978).
43. H. Heukelekian and A. Heller, Relations between food concentration and surface bacterial growth, *Journal of Bacteriology*, 40, 547 (1940).
44. H. J. Conn and J. E. Conn, The stimulating effect of colloids upon the growth of certain bacteria, *Journal of Bacteriology*, 39, 99 (1940).
45. J. H. Harwood and S. J. Pirt, Quantitative aspects of growth of the methane oxidizing bacterium *Methylococcus capsulatus* on methane in shake flask and continuous chemostat culture, *Journal of App!ied Bacteriology*, 35, 597 (1972).
46. G. Stotzky, Influence of clay minerals on microorganisms. II. Effect of various clay species, homionic clays, and other particles on bacteria, *Canadian Journal of Microbiology*, 12, 831 (1966).
47. G. Stotzky and L. T. Rem, Influence of clay minerals on microorganisms. I. Montmorillonite and kaolinite on bacteria, *Canadian Journal. Microbiology*, 12, 547 (1966).
48. D. L. King and R. D. Verma, The role of particulate substances in biotic degradation of organic waste, *Proceedings of the 23rd Purdue Industrial Waste Conference*, 75 (1968).
49. R. W. Harvey and L. Y. Young, Enumeration of particle- bound and unattached respiring bacteria in the salt marsh environment, *Applied Environ,ental Microbiology*, 40, 1, 156 (1980).
50. M. W. LeChevallier, C. D. Cawthon, and R. G. Lee, Mechanisms of bacterial survival in chlorinated drinking water, *Proceedings of the International Conference on Water Wastewater Microbiology*, Irvine, CA, February 8 to 11 (1988).
51. T. J. Marrie and J. W. Costerton, Prolonged survival of *Serratia marcescens* in chlorhexidine, *App!ied Environmental Microbiology*, 42, 1093 (1981).
52. K. C. Marshall, Adsorption of microorganisms to soils and sediments. In: *Adsorption of Microorganisms to Surfaces*, G. Bitton and K. C. Marshall (eds.), Wiley, New York (1980).
53. G. Henry, D. Prasad, and W. Lohaza, Survival of indicator Bacteria during Leaching, presented at Joint Canadian Society of Civil Engineers-American Society of Civil Engineers *National Conference onEenvironmental Engineering*, Vancouver, July 13 to 15 (1988).
54. G. M. Black and C. Webb, An immobilization technology based on biomass support particles. In: *Process Engineering Aspects of Immobilized Cell Systems*, C. Webb, G. M. Black, and B. Atkinson (eds.), Pergamon, Elmsford, NY (1986).
55. US. EPA, *Demonstration and Evaluation of the CAPTOR Process for Sewage Treatment*, US Environmental Protection Agency # PB 89-118 665/AS, Cincinnati, OH (1989).
56. US EPA, *Project Summary: Demonstration and Evaluation of the CAPTOR Process for Sewage Treatment*, U. S. Environmental Protection Agency, # EPA/600/S2-88/060, Risk Reduction Engineering Laboratory, Cincinnati, OH, Feb. (1989).
57. P. F. Cooper, I. Walker, H. E. Crabtree, and R. P. Aldred, Evaluation of the CAPTOR process for uprating an overloaded sewage works. In: *Process Engineering Aspects of Immobilized Cell Systems*, C. Webb, G. M. Black, and B. Atkinson (eds.), Pergamon, Elmsford, NY (1986).
58. P. E. Tharp and M. Frymier, High intensity biological systems using the captivated sludge process, presented at *59th Water Pollution Control Federation Conference*, Los Angeles, October 5 to 9 (1986).
59. C. E. Tharp high rate nitriftcation with CAPTOR Process, report from studies conducted by S. K. Banerji and J. N. Lin, University of Missouri, Columbia (1988).

60. F. Rogalla and M. Payraudeau, Tertiary nitrification with fixed biomass reactors, presented at *IAWPRC Conference*, Brussels, Belgium, November 24 to 28 (1987).
61. F. Rogalla and J. Jarosz, Upgrading high load activated sludge plants with biomass support systems - comparison of porous carriers with fixed submersible beds, presented at *60th Water Pollution Control Federation Conference*, Philadelphia, October 4 to 7 (1982).
62. W. Hegemann, A combination of the activated sludge process with fixed film biomass to increase the capacity of waste water treatment plants, *Water Science and Technology*, 16, 119 (1984).
63. S. R. Richards, M. Davies, and C. Hastwell, An evaluation of the CAPTOR process: a controllable fixed film process for wastewater treatment. In: *Process Engineering Aspects of Immobilized Cell Systems*, C. Webb, G. M. Black, and B. Atkinson (eds.), Pergamon, Elmsford, NY (1986).
64. W. C. Boyle and A. T. Wallace, *Status of Porous Biomass Support Systems for Wastewater Treatment: An Innovative/Alternative Technology Assessment*, Project Summary, EPA/600/S2-86/019, Environmental Protection Agency, Washington, DC (1986).
65. US EPA, *Innovative and alternative Technology Assessment Manual*, US Environmental Protection Agency, EPA/430/9-78-009, Washington, DC (1980).
66. J. W. Smith and H. A. Kharajian, Activated fixed film biosystems in wastewater treatment, *Proceedings of First International Conference on Fixed-Film Biological Processes*, Kings Island, OH, April 20–23 (1982).
67. J. Park, S. Takizawa, H. Katayama, and S. Ohgaki, Biofilter pretreatment for the control of microfiltration membrane fouling, *Water Supply*, 2, 2, 193 (2002).
68. Bohn Biofilter Corp., *What is Biofiltration*, WWW.bohnbiofilter.com/html/What_is_Biofiltration_.html, (2004).
69. Water Online, *Wastewater Biofilter*, www.wateronline.com/content/productshowcase/product.asp? (2004).
70. Waterloo Biofilter Systems, *The Future of On-Site Wastewater Treatment and Disposal*, www.waterloo-biofilter.com, (2004).
71. N. K. Shammas, Wastewater management and reuse in housing projects, *Water Reuse Symposium IV, Implementing Water Reuse*, pp. 1363–1378, AWWA Research Foundation, Denver, CO, August 2–7 (1987).
72. N. K. Shammas, An allosteric kinetic model for the nitrification process, *Proceedings of the Tenth Annual Conference of Water Supply Improvement Association*, pp. 1–30, Honolulu, HI, July (1982).
73. Hunter Water, *Burwood Beach Wastewater Treatment Works*, PDF File,Hunter Water Web Site, June 20, www.hunterwater.com.au/docs/reports/Burwood%20WWTW.pdf (2002).
74. Metcalf and Eddy, *Wastewater Engineering Treatment and Reuse*, 4th ed., McGraw Hill, New York (2003).
75. A. Vesilind, *Wastewater Treatment Plant Design*, Water Environment Federation and IWA Publishing, Alexandria, VA, USA (2003).
76. NSFC, *Technical Evaluation of the Vertical Loop Reactor Process Technology*, U. S. EPA Project No. WWPCRE13, Office of Water, National Small Flows Clearinghouse, Morgantown, WV, Sept. (1992).
77. J. M. Smith & Associates, *Technical Evaluation of the Vertical Loop Reactor Process Technology*, US Environmental Protection Agency, November (1991).
78. R. A. Brandt, E. J. Brown, and G. B. Shaw, Innovative Retrofit without Federal Funds: Brookville, Ohio Wastewater Treatment Facilities, *63rd Annual Meeting of the Ohio Wastewater Pollution Control Association*, June 16, (1989).

79. Telephone conversations and correspondence with George Smith of *Envirex* and miscellaneous information provided by *Envirex* regarding design criteria, budget costs, etc. (1991).
80. G. L. Huibrestse, C. W. Smith, D. J. Thiel, and J. W. Wittmann, *Introduction to the Vertical Loop Reactor Process*, June 12 (1986).
81. City of Willard, *Waste Water Treatment Plant*, Web Page www.willardohio.com/wwtp.htm (2004).
82. US Filter, *Envirex Products, Wastewater Treatment-Biological Treatment*, Web Page address www.usfilterenvirex.com/products/wastewater/biological.html (2004).
83. U.S. ACE, *Civil Works Construction Cost Index System Manual*, US Army Corps of Engineers, Washington, DC, PP 44, (2008-Tables) (2008).

APENDIX

United States yearly average Cost Index for Utilities (83)

Year	Index	Year	Index
1967	100	1988	369.35
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

17
Emerging Attached-Growth Biological Processes

Nazih K. Shammas and Lawrence K. Wang

CONTENTS

Abstract Among the emerging attached-growth biological treatment processes covered in this chapter are fluidized bed reactor (FBR), packed bed reactor (PBR), biological aerated filter (BAF), and hybrid biological-activated carbon systems including downflow conventional biological GAC systems and upflow fluidized bed biological GAC system (FBB-GAC). This chapter describes the above processes and explains their practice, limitations, process design, performance, energy requirements, process equipment, energy requirements, costs and case studies.

Key Words Attached-growth processes• BAF and hybrid systems (FBB-GAC)• FBR• PBR.

1. FLUIDIZED BED REACTORS (FBR)

Fluidized-bed reactors (FBR), packed-bed reactors (PBR), and biological aerated filters (BAF) represent attached growth processes that have been used to some extent for nitrification of municipal wastewaters. Unlike trickling filters, the hydraulic design of these systems is such that the media are submerged in the reactor liquid. In packed-bed reactors and biological aerated filters, the media are stationary during normal operation, held in place by gravity. In the fluidized-bed reactor, the media are expanded or fluidized as the incoming flow passes upward through the reactor.

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_17

1.1. FBR Process Description

In the conventional biological fluidized-bed reactor, often referred to as an expanded-bed reactor, wastewater or wastewater plus recycled effluent is introduced at the bottom of the reactor at a hydraulic loading rate or upflow velocity sufficient to expand the bed media, resulting in a fluidized state. The fluidized media particles provide a vast surface area for biological growth, in part leading to the development of a biomass concentration approximately five to ten times greater than that normally maintained in a conventional suspended growth reactor (1). To date, the media employed in most full-scale fluidized-bed reactors have either been silica sand or granular activated carbon.

The mechanical components and subsystems critical to the development of fluidized-bed commercial systems are (2):

(a) The device or method to distribute the influent flow to the reactor.
(b) The device or method to transfer oxygen in a controlled fashion to the fluidized-bed reactor in aerobic applications of the technology. The oxygenation system is particularly critical in the treatment of wastewaters containing medium to high concentrations of oxygen demanding material (i.e., O_2 requirements greater than 25 mg/L).
(c) The device or method to control the expansion of the fluidized bed because of biofilm growth. The bed height control system is particularly critical in treatment applications where the net yield of biomass is significant. Further details concerning the critical components have been presented elsewhere (3).

Although the development of water and wastewater systems using a fluidized bed of biomass can be traced back to the 1940s in England (4), media-based fluidized-bed reactors were not developed until the early 1970s. Researchers at Manhattan College in New York, at the EPA Municipal Environmental Research Laboratory in Cincinnati, OH, and at the Water Research Centre in Medmenham, England, can be credited for the initial application of media-based fluidized-bed reactors to water and wastewater treatment. The Manhattan College researchers were granted a US patent in 1974 (assigned to Ecolotrol, Inc.) for the application of the fluidized-bed process configuration to "denitrifying wastewater" (5). In a paper published in 1970 by researchers from the University of Michigan, biological activity was observed in expanded-bed activated carbon reactors and was believed to be the reason for the observed nitrate reduction (6).

The ability of the biological fluidized-bed process configuration to intensify biological reaction rates through accumulation of high concentrations of active biomass has attracted attention for many years (7). The results from laboratory and field pilot scale studies have consistently illustrated the technical advantages of the fluidized bed over most other suspended and attached growth reactor configurations in many wastewater treatment applications. In 1981, a comprehensive account of ongoing fluidized-bed process development activities was published based on a 1980 seminar held in Manchester, England (8). Although hailed at that time as the most significant development in the wastewater treatment field in the last 50 years, it also was claimed that no full-scale plants were yet in operation. Since that time, even though more than 70 commercial, fluidized-bed reactors have been installed in North America and Europe, wider use of the technology has been hampered by such factors as (9):

(a) Mechanical scale-up issues.
(b) Slow development of economically attractive system configurations.
(c) Proprietary constraints.

According to a 1991 state-of-the-art review of fluidized beds for water and wastewater treatment, the technology was being applied largely for industrial versus municipal wastewater treatment at current operating full-scale installations in North America and Europe (9). Although full-scale fluidized-bed industrial systems are operating under conditions that result in nitrification (10), few, if any, systems have been installed for nitrification of municipal wastewaters on a full scale. A limited number of reactors have been installed for denitrification of municipal wastewater (11).

1.2. Process Design

Information useful for the process design of full-scale systems for nitrification of municipal wastewater derived from the results of fluidized-bed pilot plant studies (12–24) is summarized as follows:

(a) A half-order model appears appropriate to describe the kinetics of ammonium oxidation in fluidized-bed reactors under nonlimiting DO conditions.
(b) The volumetric removal rate and the specific ammonium oxidation rate decrease significantly at low reactor ammonium concentrations.
(c) The fluidized-bed hydraulic retention time required to achieve nitrification down to ammonium levels of 2 mg/L or less ranges from 10 to 40 min. This HRT is for treatment of municipa1 wastewaters containing less than 50 mg/L of $CBOD_5$ and approximately 20 mg/L of oxidizable nitrogen compounds, and providing that the reactor is designed to promote the buildup of at least 8.5 g/L of volatile attached solids and that nonlimiting DO conditions are achieved. The actual HRT required will depend on such factors as the concentration of carbonaceous BOD in the wastewater, the system hydraulics (i.e., plug flow versus complete mixing conditions), and the reactor temperature and pH conditions.

If the use of the fluidized bed for nitrification is being considered, onsite piloting is recommended given the limited amount of full-scale operating and performance information on this application.

1.3. Applications

The fluidized-bed reactor is more commonly used for industrial wastewater rather than municipal wastewater. Concerns over municipal applications have included mechanical scale-up factors, proprietary constraints, and economically unattractive system appurtenances (25). However, there are successful municipal applications; Table 17.1 lists the design parameters and loadings of four industrial and municipal installations with fluidized-bed reactors operating in the denitrification mode (1).

The principal commercial suppliers of fluidized-bed systems are Dorr-Oliver, Envirex, and Ecolotrol. Both Dorr-Oliver and Envirex systems were developed on the basis of Ecolotrol process patents. Currently, Envirex is the only manufacturer actively marketing the fluidized-bed reactor for denitrification applications in the United States. Table 17.2 summarizes the types of reactors in use (1).

Table 17.1
Design parameters and loadings of denitrification FBR plants (2)

Parameter[a]	Facility			
	Pensacola[b,c]	Reno-Sparks	Rancho, CA[c]	IBM[c]
Mean wastewater flow, L/s	1052	1883	263	113
Mean wastewater flow, MGD	24	43	6[d]	1[d]
Maximum wastewater flow, L/s	1490	2400		
Maximum wastewater flow, MGD	34	55		
Influent NO_3^--N, mg/L	20	18	21	54
Effluent NO_3^--N, mg/L	< 6	2	2.5	8
Design wastewater temperature, °C	18	13	22	10
Estimated reactor biomass, mg/L VSS	NA	18,000	28,000	NA
Hydraulic retention time[e], min	8.5	13.8	10	26
Hydraulic loading rate[f], $m^3/m^2/day$	672	550	336	578
Hydraulic loading rate[f], gpd/ft^2	11.4	9.3	0.8	1.3
Estimated settled sand depth, m	1.8	2.4	1.2	1.5
Estimated settled sand depth, ft	6	8	4	5
Fluidized-bed height, m	4	4.9	2.4	2.7
Fluidized-bed height, ft	13	16	8	9

[a] All values are yearly averages.
[b] Modified design as developed by Dorr-Oliver, Inc.
[c] No longer operated for denitrification.
[d] Equalization provided to achieve a constant wastewater flow rate.
[e] Based on mean wastewater flow and fluidized-bed/empty-bed volume.
[f] Based on total flow to the reactor (plant flow plus recycle).

Table 17.2
Types of FBRs in use (2)

Oxitron System
- – Developed by Dorr-Oliver
- – System based on Ecolotrol process patents
- – Uncertain regarding system marketing in North America
- – Dorr-Oliver Europe marketing systems in Europe

Rex aerobic fluidized-bed process, anaerobic and biological denitrification configuration
- – Developed by Envirex/Ecolotrol based on Ecolotrol process patents
- – Sold in North America by Envirex

Custom engineered systems
- – Developed by consulting engineering firms
- – Normally designed and operated under conditions falling outside the limits of Ecolotrol patents

The principle of the fluidized-bed reactor is the same, regardless of the application. Examples of applications to the remediation of groundwater to remove various organic contaminants and produce cleaner and safer water supplies can be found in the literature (26–31).

1.4. Design Considerations

The upflow fluidized-bed system usually consists of a reactor vessel in the form of an above-ground steel and fiberglass tower or in-ground concrete reactors. The flow rate and strength of waste determines the size of the reactor vessel. The reactor size is dependent on temperature; at 15 °C (59 °F), the design loading rate is 6, 420 kg NO_3^--N/1, 000 m^3/day (400 lb/1, 000 ft^3/day) (27).

When the fluidized bed system is operated for denitrification, methanol is fed to the nitrified influent by injection into the recycle line (see Fig. 17.1). The reactor operates as a plug flow process; however, the high recycle ratio of reactor effluent to plant flow (10:1 to 20:1 for high strength waste treatment and 2:1 to 5:1 for municipal denitrification) emulates a complete mix system. The high recycle ratio also helps protect the reactor from shock loads and is required to achieve bed fluidization. The amount of recycle is dictated by a maximum allowable fluid-bed height, structural considerations often control bed height (33–36).

1.5. Case Study: Reno-Sparks WWTP

A flow diagram for the 1,753-L/s (40-MGD) Reno-Sparks Wastewater Treatment Plant is shown in Fig. 17.2. The treatment plant, which serves the cities of Reno and Sparks in Nevada, consists of preliminary treatment, primary treatment, phosphorus and BOD removal in a sidestream phosphorous-removal system, nitrification biotowers, denitrification upflow fluidized-bed reactors, post aeration, effluent filtration and disinfection. The solids-handling system consists of thickening, anaerobic digestion, and dewatering.

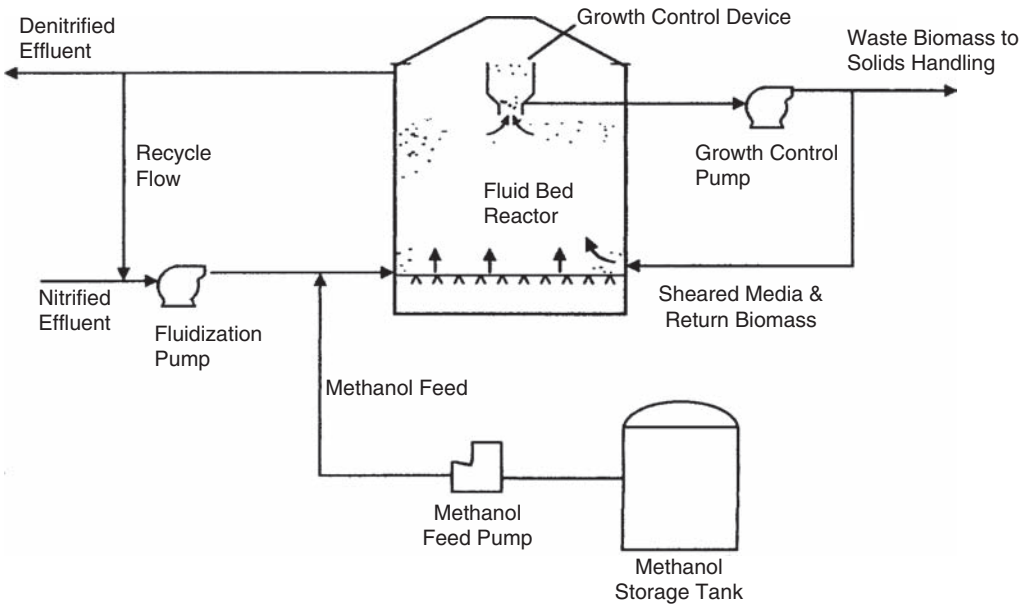


Fig. 17.1. Flow diagram of Upflow Fluidized-Bed System (2).

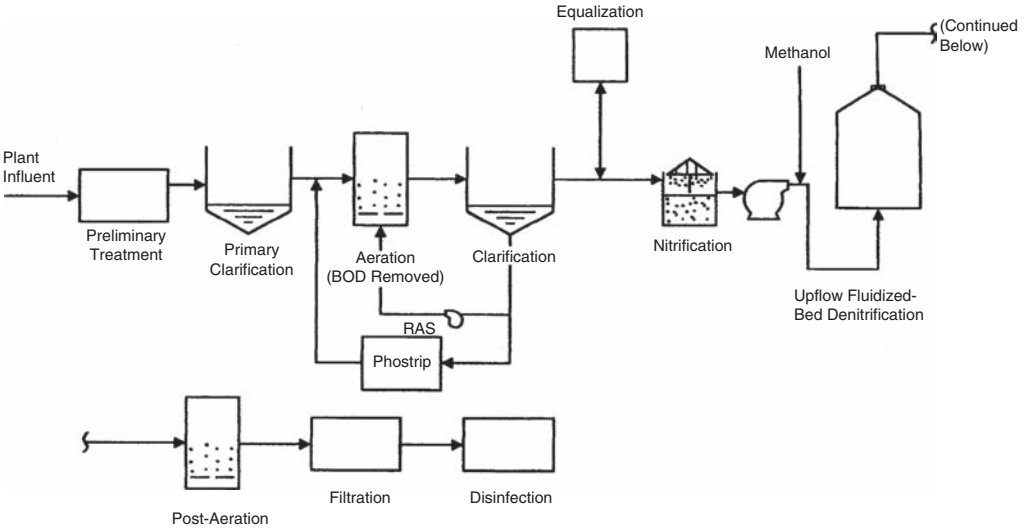


Fig. 17.2. Flow diagram of Reno-Sparks Wastewater Treatment Plant (2).

The denitrification system consists of four upflow fluidized bed towers measuring approximately 8.2 m (27 ft) in diameter by 6.2 m (20.5 ft) high. The hydraulic residence time at average daily flow is 13.8 min, and the solids residence time (SRT) is 8.5 days. The denitrification system, manufactured by Envirex, was designed to produce effluent with a nitrate level of 2 mg/L. A summary of monthly plant operating data is provided in Table 17.3. The data indicate that the Reno-Sparks plant has consistently met its effluent requirements, with an average effluent ammonia level of 0.16 mg/L and a NO_x-N level of 0.29 mg/L. The plant's efficiency in removing total nitrogen has been 94%. The removal rate of the fluidized-bed reactors has been 6.4 kg NO_x-N/m^2/day (1.3 lb/ft^2/day), and the plant has regularly produced an effluent TN of less than 3 mg/L and an average effluent TN of 1.78 mg/L. The one event over 3 mg/L TN was 3.55 mg/L.

2. PACKED BED REACTOR (PBR)

2.1. Aerobic PBR

A packed-bed reactor, often referred to as a submerged filter, contains a stationary bed of media that provides support for biological growth. The influent wastewater (or wastewater plus recycled effluent) is normally introduced at the bottom of the reactor through a flow distribution system. Methods utilized to supply the necessary oxygen to support biomass growth have included direct introduction of air (37) or high-purity oxygen (38) into the bottom of the reactor through a gas distribution system or injection of air or oxygen into the feed line entering the reactor. Alternatively high-purity oxygen has been dissolved in the feed stream in an oxygenation device before the feed entering the reactor (37).

Table 17.3
Reno-Sparks Wastewater Treatment Plant: Monthly performance data (2)

	Plant influent					Denitrification process influent		Denitrification process effluent		Final plant effluent				
	Flow		BOD_5	NH_4^+-N	TN	NH_4^+-N	NO_x-N	TSS	NO_x-N	BOD_5	TSS	NH_4^+-N	NO_x-N	TN
Month	L/s	MGD	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L
JAN 90	1133	25.86	167			0.37	14.44		2.12	5	5	0.13	0.79	2.53
FEB 90	1153	26.30	171			0.53	14.58		3.91	8	5	0.24	0.43	2.25
MAR 90	1174	26.79	178			0.31	14.80		0.80	5	4	0.27	0.88	2.79
APR 90	1156	26.37	177			1.37	14.21		0.82	6	2	0.29	0.90	3.55
MAY 90	1169	26.66	183			0.67	12.84		0.19	5	3	0.08	0.15	1.38
JUN 90	1221	27.87	175			0.35	14.02		0.06	3	2	0.11	0.16	1.43
JUL 90	1202	27.42	152			0.30	14.07		0.04	3	2	0.10	0.06	1.11
AUG 90	1233	28.14	164			0.91	13.70		0.29	6	10	0.32	0.14	1.82
SEP 90	1211	27.64	168			0.56	14.18		0.12	7	18	0.41	0.11	2.50
OCT 90	1175	26.81	149			0.15	11.03		0.03	3	2	0.05	0.01	1.01
NOV 90	1144	26.10	162			0.37	13.17		0.10	4	3	0.19	0.07	1.26
DEC 90	1192	27.20	150			1.13	13.92		0.22	4	6	0.33	0.37	2.00
JAN 91	1168	26.64	132	21.1	32.3	0.45	14.04	22	0.24	3	4	0.17	0.35	1.64
FEB 91	1122	25.60	130	22.2	32.3	0.46	14.26	20	0.16	3	3	0.18	0.28	1.58
MAR 91	1149	26.21	135	22.5	32.2	0.35	14.88	18	0.39	3	4	0.10	0.42	1.70
APR 91	1103	25.17	142	22.7	32.2	0.16	15.56	20	0.31	4	4	0.02	0.28	1.53
MAY 91	1137	25.95	186	22.0	31.9	0.59	15.07	21	0.17	4	4	0.03	0.14	1.47
JUN 91	1171	26.72	167	21.0	31.9	0.08	13.72	17	0.03	2	4	0.02	0.03	1.18
JUL 91	1210	27.36	160	20.1	31.6	0.06	13.85	13	0.08	3	4	0.03	0.03	1.01
AVERAGE	1169	26.67	160	21.7	32.1	0.48	14.02	19	0.53	4	5	0.16	0.29	1.78

In 1975, the EPA Process Design Manual for Nitrogen Control noted that packed-bed reactors for nitrification were a comparatively recent development, having progressed from laboratory and pilot status to the point of commercial availability (39). Since that time packed-bed reactors have been widely applied for commercial treatment of industrial wastewaters and contaminated ground waters. Despite continuing interest in packed-bed reactors for nitrification of municipal wastewaters (37–42) and additional pilot studies, packed-bed reactors have not been widely applied on a full scale. The lack of information clearly demonstrating significant advantages of the technology relative to alternatives for this application has limited the acceptance of packed-bed reactors at the full-scale level for municipal wastewater treatment.

Several types of media including stones, gravel, anthracite, and random plastic media had been successfully utilized in pilot plant studies of packed-bed reactors. In more recent studies, the media utilized has normally been either random or corrugated plastic structures with high void volume (37–42). The use of such media may eliminate the need for backwashing to control the buildup of reactor SS. If solids buildup is not prevented or controlled, the hydraulic integrity of the reactor will be compromised. Design and operating strategies that minimize the buildup of reactor SS Include:

(a) The use of media with a high void volume (greater than 90%)
(b) The supply of oxygen by the direct introduction of air into the bottom of the reactor

2.2. *Anaerobic Denitrification PBR*

2.2.1. Coarse Media Beds

When PBRs are used for denitrification, nitrates and nitrites are reduced to nitrogen gas through the action of facultative heterotrophic bacteria. Coarse media denitrification filters are attached growth biological processes in which nitrified wastewater is passed through submerged beds containing natural (gravel or stone), granular activated carbon (GAC) or synthetic (plastic) media. The systems may be pressure or gravity. Minimum bed media size is about 15 mm. Anaerobic or near anaerobic condition is maintained in the submerged bed, and because the nitrified wastewater is usually deficient in carbonaceous materials a supplemental carbon source (usually methanol) is required (Fig. 17.3) to maintain the attached denitrifying slime (43). Because of the high void percent and low specific surface area characteristic of high porosity coarse denitrification filters, biomass (attached slime) continuously sloughs off. As a result, the coarse media column effluent is usually moderately high in suspended solids (20 to 40 mg/L), requiring a final polishing step.

A wide variety of media types may be used as long as high void volume and low specific volume are maintained. Both dumped plastic media (Fig. 17.4) and corrugated sheet media have been used. Backwashing is infrequent and is usually done to control effluent suspended solids rather than pressure drop. Alternate energy sources such as sugars, volatile acids, ethanol, or other organic compounds, as well as nitrogen deficient materials such as brewery wastes may be used. Nitrogen gas filled coarse media denitrification filters are a possible modification.

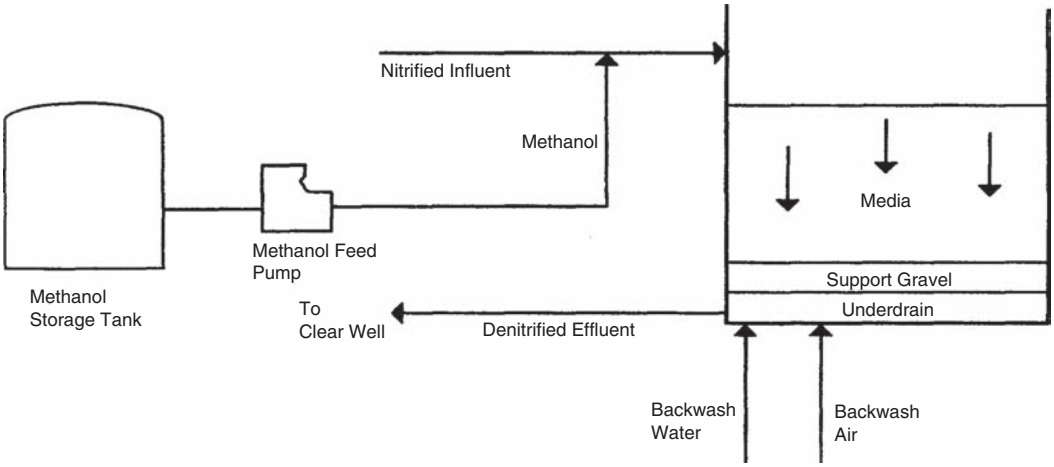


Fig. 17.3. Flow diagram of Packed Bed Reactor System (2).

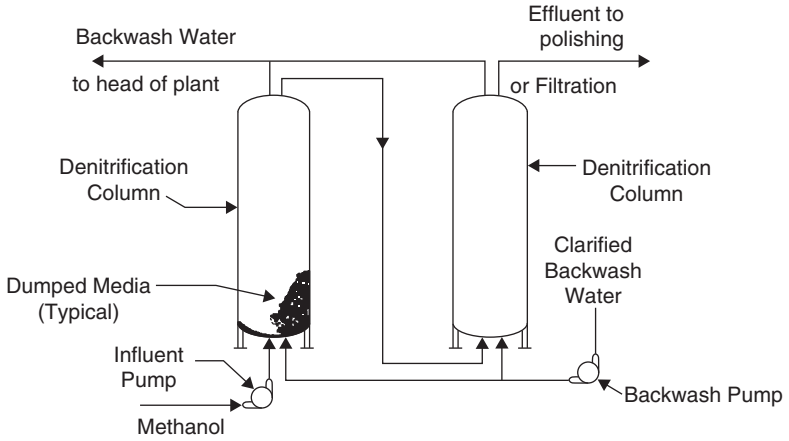


Fig. 17.4. PBR System with coarse media denitrification columns (43).

2.2.2. Fine Media Beds

The fine media denitrification filter is an attached growth biological process in which nitrified wastewater is passed through a pressurized submerged bed of sand or other fine filter media (up to about 15 mm in diameter) in which anoxic conditions are maintained. Because of the relatively fine media used, physical filtration analogous to that occurring in a pressure filter takes place. As a result, a clear effluent is produced, eliminating the need for final clarification (43). Backwashing is required to maintain an acceptable pressure drop. Surface loading rates may be somewhat lower than those common for pressure filtration. Development of the denitrifying slime and consequent denitrification efficiency are a function of the specific surface area of the filter, and in practice fine media denitrification filters convert nitrates to

nitrogen gas at a much higher rate than suspended growth systems. The coarser the media, the less frequent the backwashing, although the effluent may be more turbid.

Common modifications include the use of various media such as garnet sand, silica sand, anthracite coal or activated carbon with varying size distributions. Multimedia systems have also been used. Alternate energy sources such as sugars, volatile acids, ethanol or other organic compounds as well as nitrogen deficient materials such as brewery wastewater may be used. An air scour may be incorporated into the backwashing cycle; however, temporary inhibition of denitrification may result. Various types of underdrains may be used. A bumping procedure (short periodic flow reversals) has been used to remove entrapped nitrogen gas bubbles produced during denitrification. Denitrification may be combined with refractory organic removal. Upflow systems utilizing fine media (sand or activated carbon) have been operated as fluidized bed reactors.

2.3. *Applications*

PBRs are used mostly for nitrogen removal by biological nitrification-denitrification of municipal wastewater that has undergone carbon oxidation. Examples of packed-bed denitrification treatment plants are listed in Table 17.4. Similar units are also used to reduce nitrate in industrial wastewater systems (43).

2.4. *Design Criteria*

The design criteria for both coarse and fine media PBRs, as stated in EPA manuals (39, 43), are given below.

2.4.1. *Coarse Media Beds*

(a) Optimum pH = 6.5 to 7.5
(b) Voids = 70% to 96%
(c) Specific surface = 65 to 274 ft^2/ft^3
(d) Media size = greater than 15 mm
(e) Loading rate in lb $NO_3^- - N/ft^2$ packing surface/day is a function of temperature up to 0.5×10^{-4} at 5 °C, 0.2 to 0.8×10^{-4} at 15 °C, and 0.8 to 1.3×10^{-4} at 25 °C
(f) Surface loading rate = 2.5 gal/ft^2/day for a flow of 0.3 MGD and 4.1 gal/ft^2/day for a flow 0.5 MGD
(g) Amount of the most common energy source, methanol, required may be estimated at 2.47 mg/L CH_3OH per mg/L of $NO_3^- - N$ and 1.53 mg/L CH_3OH per mg/L $NO_2^- - N$ in the inlet to the process

2.4.2. *Fine Media Beds*

(a) Flow Scheme: Downflow (although upflow systems with different design criteria have been utilized).
(b) Optimum pH = 6.5 to 7.5.
(c) Voids = 40% to 50%.
(d) Specific surface = 85 to 300 ft^2/ft^3.
(e) Media diameter (d_{50}) = 2 to 15 mm.
(f) Surface loading Rates = 0.5 to 7.0 gpm/ft^2.
(g) Column depth = 3 to 20 ft (function of specific surface and contact time).

Table 17.4
Installation list of packed bed denitrification systems (2)

Facility and location	Description of packed-bed denitrification system	Capacity	
		L/s	MGD
Tampa, Florida Hookers Point AWTP	Twelve 97 m^2 (1,050 ft^2) filters Nineteen 93 m^2 (1,000 ft^2) filters	4,208	96.0
Seminole County, Florida NW Area Regional WW Facility Expansion	Two 46 m^2 (500 ft^2) filters	110	2.5
Port Orange, Florida	Six 52 m^2 (560 ft^2) filters	351	8.0
Hillsborough County, Florida Valrico Wastewater Facility	Three 46 m^2 (500 ft^2) filters	132	3.0
U.S. Home Brandon, Florida	One 19 m^2 (200 ft^2) filter	33	0.75
Purity Farms Clearwater, Florida	One 9.3 m^2 (100 ft^2) filter	10	0.23
Hillsborough County, Florida Dale Mabry AWTP	Five 60 m^2 (650 ft^2) filters	264	6.0
Piney Orchards, Maryland	Four 9.3 m^2 (100 ft^2) filters	53	1.2
Hillsborough County, Florida Falkenburg RD AWTP	Five 46 m^2 (500 ft^2) filters	264	6.0
Altamonte Springs, Florida	Seven 56 m^2 (600 ft^2) deep-bed filters for tertiary filtration, denitrification, and virus control of municipal sewage treatment plant effluent	110 548	2.5 (Avg.) 12.5 (Peak)
Florida Cities Water Co. Flesta Village Fort Myers, Florida	Four 37 m^2 (400 ft^2) filters for nitrate reduction and SS removal	96 220	2.2 (Avg.) 5.0 (Peak)
Kanapaha Wastewater Treatment Plant Gainesville, Florida	Six 46 m^2 (500 ft^2) filters	308 770	7.0 (Avg.) 17.5 (Peak)
Parkland Ill Expansion Islip, New York	Deep-bed gravity denitrification-effluent polishing system including four 5.6 m^2 (60 ft^2) filters	11	0.26
Fairfield Village New York	Two 5.6 m^2 (60 ft^2) deep-bed sand filters for effluent polishing and denitrification	4	0.085
Southhampton Hospital Southhampton, New York	Two 4.7 m^2 (50 ft^2) deep-bed sand filters for effluent polishing and denitrification	4	0.1

(Continued)

Table 17.4
(Continued)

Facility and location	Description of packed-bed denitrification system	Capacity L/s	Capacity MGD
Blue Ridge Condo, Medford	One deep-bed sand filter system. System includes three deep-bed gravity filter cells 5.6 m^2 (60 ft^2) each	9	0.2
Brookhaven, New York			
Parkland III	One deep-bed gravity filtration system for effluent polishing and denitrification. System includes four deep-bed filter cells 5.6 m^2 (60 ft^2) each	28	0.65
Islip, New York			
Parr Village	Three 4.8 m^2 (52 ft^2) deep-bed sand filters	20	0.45
Yaphank, New York			

(h) Backwash rate = 8 to 25 gpm/ft^2.
(i) Backwash cycle frequency = 0.5 to 4.0 days.
(j) Amount of the most common energy source, methanol, required may be estimated at 2.47 mg/L CH_3OH per mg/L of NO_3^- − N and 1.53 mg/L CH_3OH per mg/L NO_2^- − N in the inlet to the process.

2.5. Performance

As with trickling filters, the efficiency and performance of nitrifying packed-bed reactors can be expected to correlate to the effective surface area for biofilm growth, although growth of active nitrifiers in the voids of the media may affect this correlation. Thus, both the surface loading and the volumetric loading are likely to influence nitrification efficiency and performance in packed-bed reactors. Other factors such as the concentration of DO, $CBOD_5$, and ammonium in the reactor, environmental conditions (i.e., temperature and pH), and media characteristics (i.e., surface-to-volume ratio and percent voids) will influence the correlations between loading and nitrification performance. Although surface and volumetric loading information applicable to the design of packed-bed reactors for nitrification of municipal wastewaters is available (37–41), onsite piloting is recommended if the technology is being considered for use on a full scale.

Packed bed reactors are capable of converting nearly all nitrates in a nitrified secondary effluent to gaseous nitrogen. Overall nitrogen removals of 70% to 90% are achievable. In fine media beds Suspended solids removals of up to 93% have been achieved. Under controlled pH, temperature, loading and chemical feed high levels of reliability are achievable. Studies on the effects of environmental factors, modeling and kinetics in full scale submerged denitrification PBRs can be found in Refs. (44) and (45).

With controlled supplemental carbon feed rates, little excess sludge is generated. Sludge production varies between 0.6 to 0.8 lb/lb NH_3 − N reduced.

2.6. Case Study: Hookers Point WWTP (Tampa Florida)

Operating data for downflow packed-bed systems are shown in Table 17.5. The 4208-L/s (96 MGD) Hookers Point Wastewater Treatment Plant (WWTP) includes preliminary treatment, primary treatment, biological treatment, post aeration, and effluent disinfection. The biological treatment system includes two-stage carbonaceous oxidation/nitrification using high-purity oxygen and a separate-stage downflow packed-bed denitrification system with methanol feed. A flow diagram is shown in Fig. 17.5.

The downflow packed-bed denitrification system consists of 20 filters measuring 3 × 32 m (10 × 105 ft). Each filter is filled with 142 cm (56 in.) of coarse sand (2.3 mm), loaded at an average rate of 59 to 117 $m^3/m^2/day$ (1 to 2 gpm/ft^2) and having an empty-bed contact time of 45 min at average flow.

The Hookers Point WWTP receives domestic wastewater, with a 30% contribution from breweries (46). The influent wastewater has a BOD_5 of 224 mg/L, TSS of 221 mg/L, and TKN of 32 mg/L. The current effluent limits of the plant are 5 mg/L for BOD_5 and TSS, 3 mg/L for TN on an annual average basis, and 7.5 mg/L for total phosphorous (TP). The average month's effluent is below 3 mg/L TN 83% of the time, with an average over 3-year period of 2.33 mg/L. It should be noted that the effluent limit was changed to 3 mg/L TN in 1990. Before that time, the limit was 4 mg/L TN in summer and 5 mg/L in winter. The average effluent TSS is 2 mg/L and is relatively stable. Hookers Point has a process loading rate of 1.32 kg NO_x-$N/m^2/day$ (0.27 $lb/ft^2/day$). The brewery waste may contribute significantly to the background nitrogen removal by synthesis. The plant's overall efficiency in removing nitrogen and SS has been 93% and 99%, respectively.

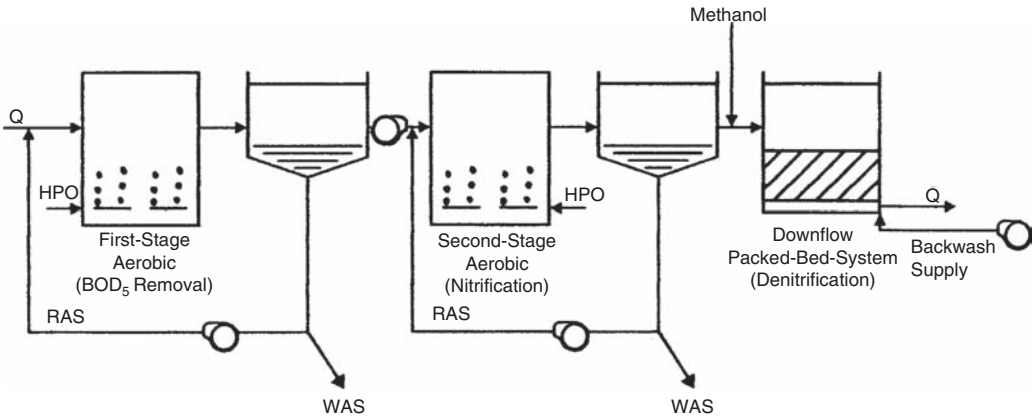


Fig. 17.5. Flow diagram of Hookers Point Advanced Wastewater Treatment Plant (2).

Table 17.5
Operating data for downflow packed bed reactor systems (2)

Facility	Florida Location	Capacity		Average Rate		Number of Denit. Filters	Filter Size	Media Depth		Media Size, mm
		L/s	MGD	$m^2/m^2/day$	gpm/ft^2			m	ln	
Hookers Point	Tampa	4,208	96.0	123	2.1	20	3 × 32 m (10 × 105 ft)	1.47	54	2.3
Fiesta Village	Ft. Myers	220	5.0	117	2.0	4	3 × 13.4 m (10 × 44 ft)	1.83	72	3.0
Altamonte Springs	Altamonte	548	12.5	123	2.1	7	3 × 18.3 m (10 × 60 ft)	1.83	72	Dual Media
Faulkensand Road, Hillsborough Co.	Tampa	264	6.0	29	0.5	5	3 × 15.2 m (10 × 50 ft)	1.22	48	3.0
Dale Mabry	Tampa	264	6.0	123	2.1	5	3 × 19.8 m (10 × 65 ft)	1.83	72	2.3
Port Orange	Port Orange	526	12.0	123	2.1	7	3 × 17.1 m (10 × 56 ft)	1.07	42	1.8

All plants have a 3 mg/L TN permit limit.

2.7. Energy Requirement

2.7.1. Coarse Media Beds

Pumping energy can be computed from the following Eq. (43):
kWh/year = (1140 MGD × ft of total average head)/wire to water efficiency

For a 0.5 MGD plant treating 14 mg/L of $NO_3^- - N$ two 10-ft diameter by 10-ft deep tanks would be required. Therefore, using 15 ft of total head and a wire to water efficiency of 0.60, 14,250 kWh will be required for wastewater pumping.

Backwashing at a rate of 20 gpm/ft^2, once a month for four hours would require an additional energy consumption of 1425 kWh/year.

Upflow and downflow operations consume roughly the same amount of energy.

2.7.2. Fine Media Beds

The energy requirement for PBR fine media beds is shown in Fig. 17.6. The assumptions for energy determination are as follows (47):

(a) Influent $NO_3 - N$ = 25 mg/L; effluent = 0.5 mg/L.
(b) Media size = 2 to 4 mm.
(c) Temperature is 15 °C.
(d) Methanol feed rate = 3 : 1 (CH_3OH : $NO_3^- - N$).
(e) Loading rate = 1.7 gpm/ft^2.
(f) Depth = 6 ft.
(g) Backwash for 15 min @ 25 gpm/ft^2 and 25 ft TDH once per 2 days for pressure and daily for gravity system.

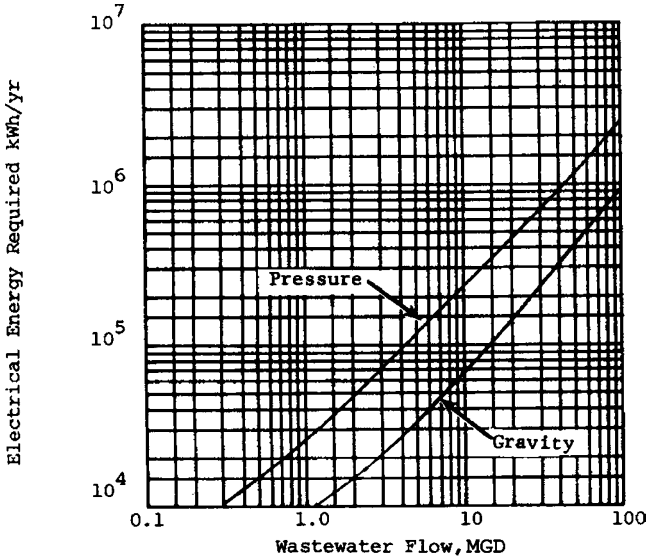


Fig. 17.6. Energy requirements for PBR System (43).

2.8. Costs

2.8.1. Coarse Media Beds

The construction cost for PBR coarse media beds is determined as follows: for a 0.5 MGD plant treating 14 mg/L $NO_3^- - N$, two 10-ft diameter by 10-ft deep tanks would be required. Construction costs (1972 dollars, Utilities Index = 141.94) for such a system was approximately USD 200,000 (39). To obtain the value in terms of the present 2008 USD, using the Cost Index for Utilities (Appendix), multiply the cost by a factor of 552.16/141.94 = 3.89 (48). Thus, the 2008 construction cost for 0.5 MGD PBR beds would be 200,000 × 3.89 = USD778,000.

The cost for chemicals (Methanol) is USD 0.03 × 3.89 = USD 0.12/1000 gal (39, 43, 48). The O & M cost for labor is USD 0.03 × 3.89 = USD 0.12/1,000 gal. Thus the total operation and maintenance cost in terms of 2008 USD would be USD 0.24 per 1000 gal treated.

2.8.2. Fine Media Beds

Construction costs (1975 dollars, Utilities Index = 190.49) for PBR fine media beds are shown in Fig. 17.7 (43).To obtain the values in terms of the present 2008 USD, using the Cost Index for Utilities (Appendix), multiply the costs by a factor of 552.16/190.49 = 2.90 (48).

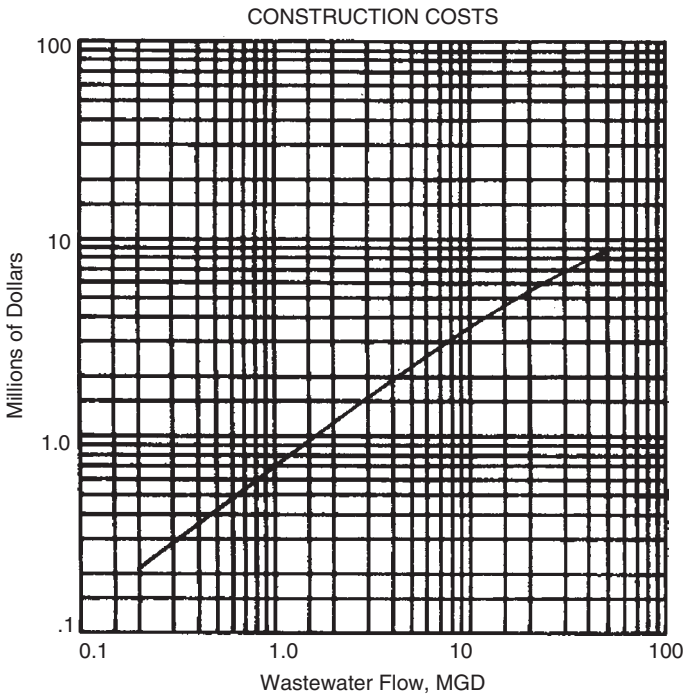


Fig. 17.7. Construction cost for PBR System (43).

The operation and maintenance costs for a 0.5 MGD plant treating 14 mg/L $NO_3^- - N$ is determined as follows: The cost for chemicals (Methanol) is USD $0.03 \times 3.89 =$ USD 0.12/1,000 gal (39, 43, 48). The O & M cost for labor (including normal maintenance and daily backwash) is USD $0.04 \times 3.89 =$ USD 0.16/1,000 gal. Thus the total operation and maintenance cost in terms of 2008 US dollars would be USD 0.28 per 1,000 gal treated.

3. BIOLOGICAL AERATED FILTER (BAF)

3.1. BAF Process Description

In the biological aerated filter (BAF), the media are submerged in the reactor and primary clarified wastewater is introduced at the top of the reactor (2). As noted in an EPA-sponsored study (49), BAF systems are very similar in both physical appearance and mode of operation to a downflow water filter or tertiary wastewater filter except that:

(a) A coarser, low density media is used.
(b) Air is diffused upward through the media during operation.

The air is introduced into the media through an air diffusion system located approximately 20 to 25 cm (8 to 10 in.) above the filter underdrain system (2, 49). This air is supplied to promote biomass growth in the voids of the packed bed and on the media surface above the air diffusion system. The function of the media below the air diffusion system is to remove SS. As newly grown biomass and influent SS build up in the reactor, the head loss across the unit increases. The unit is backwashed when a predetermined headloss is reached. The backwashing operation involves a series of air scours and liquid flushes with treated effluent. The intent of this operation is to release SS trapped in the voids of the packed bed and to control the extent of film growth on the media surface (2). The backwash water is either thickened separately or conveyed to primary clarification at the head end of the plant. A common process flow diagram for a complete Biocarbone BAF system is shown in Fig. 17.8

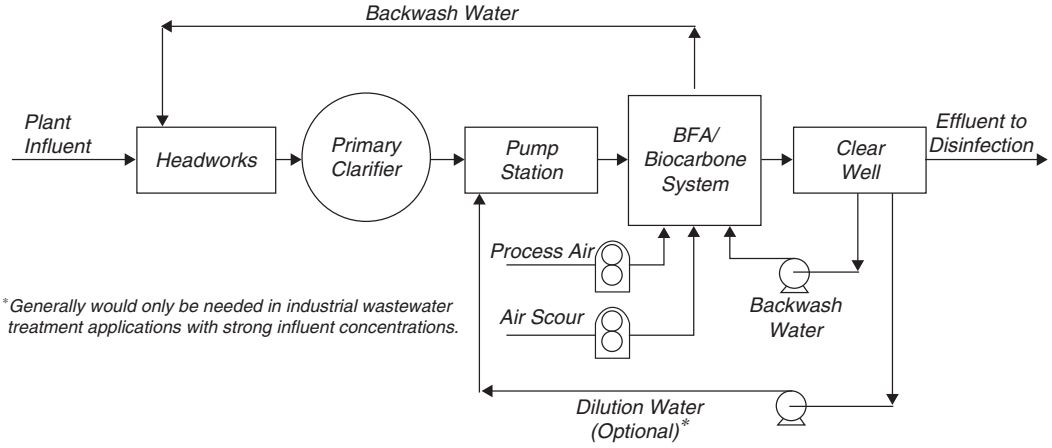


Fig. 17.8. Flow diagram for a Biological Aerated Filter (BAF) System (50).

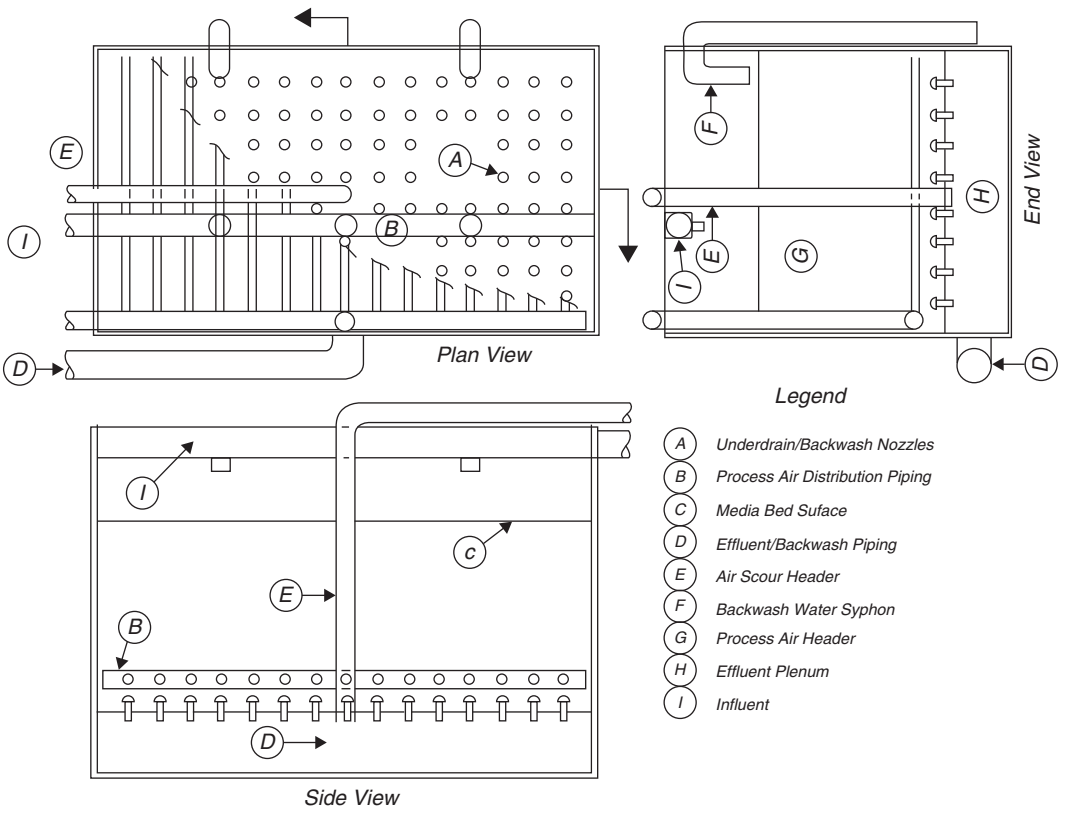


Fig. 17.9. Plan and side views of a BAF/Bicarbone unit (50).

and details of a Bicarbone filter unit is shown in Fig. 17.9. Biocarbone is the trademark name given to Omnium de Traitement et de Valorisation (OTV) commercial embodiment of the process.

When treating primary effluent, the BAF/Biocarbone process can be designed to achieve carbonaceous BOD removal only or carbonaceous BOD removal and nitrification by selecting appropriate loading rates. The process can also be designed to achieve advanced secondary treatment removals of BOD and suspended solids as well as nitrification with either primary or secondary effluent feed (50).

The primary advantage of the BAF is biological treatment and solids separation in the same reactor eliminating the requirement for separate secondary clarification. Consequently, the technology could reduce the space requirements for treatment relative to more conventional technologies such as the activated sludge system (2).

The advantages of the BAF process can be summarized as follows:

(a) Absence of secondary clarifier.
(b) Compactness. Good alternative when land availability is low or expensive because the reactor has a compact footprint.

(c) Modular design and implementation to suit various flow conditions and effluent quality requirements.
(d) Considerable inertia against pollution breakthrough under load variations with peak flows up to three times the average.
(e) The rapid start-up (relative to activated sludge) allows for adjustment in the number of units in service to match the pollution load arriving at the plant.

3.2. Applications

The first commercial, full-scale BAF system began operation in 1982 in Soissons, France (51). Since that time a number of systems have been installed in Europe, Japan, and North America (49, 52). As of 1990, there were approximately 30 commercial full-scale Biocarbone BAF systems installed or under construction, designed at wastewater flows of 22 L/s (0.5 MGD) or greater (52). The largest Biocarbone BAF system installed to date is designed to treat approximately 1,056 L/s (24 MGD) (51). Most Biocarbone BAF systems in operation today have been designed for $CBOD_5$ and TSS removal, but the systems can be designed to nitrify primary or secondary effluent.

3.3. BAF Media

The original media employed in the Biocarbone BAF was activated carbon. This material had the desirable characteristics of a porous surface with a high surface-to-volume ratio for enhancing biomass attachment and a low specific gravity to allow for ease of air scouring and backwashing, but it was found too expensive. Subsequently, alternative granular media has been used for economic reasons. The media in most currently operating BAF systems consists of a kiln-fired clay or shale particle. Biodamine and Biodagene are the names given to two of the media often used in the Biocarbone BAF (49). Biodamine is an angular shaped media whereas Biodagene is more spherical.

The angularity and size range of the media significantly affects the BAF treatment performance and operating requirements. The use of smaller media in the range of 2 to 4 mm (0.08 to 0.16 in.), although it offers a superior effluent quality to that of a system with larger sized media, normally requires more frequent backwashing (49). The smaller media has been recommended when nitrification is required (52). Expected effluent quality as a function of media gradation is shown in Table 17.6 (50).

Table 17.6
Recommended BAF media gradation (50)

Media gradation (mm)	Effluent BOD_5 (mg/L)	Effluent TSS (mg/L)
2–4	10	10
3–6	20	20
4–8	30	30

3.4. Process Design and Performance

OTV through years of conducting pilot- and full-scale Biocarbone plant evaluations has developed reliable correlations between applied pollutant and/or hydraulic loading rates and effluent quality or percent pollutant removal (49, 50).

One of these generalized correlations is depicted in Fig. 17.10 for two types of media, activated carbon and biodamjne (vitrified clay particles). Effluent quality from a Biocarbone unit is graphically depicted in Fig. 17.11.

Pilot plant studies by the developer of the Blocarbone BAF system (49) indicate that for a system treating primary effluent wastewater containing a high $CBOD_5$ concentration, nitrification is governed in part by the COD volumetric loading. The volumetric loading is based on the volume occupied by the media (i.e., empty bed volume). The results (Fig. 17.12) indicate that at a COD volumetric loading above 0.2 lb/ft^3/day (3.2 kg/m^2/day), nitrification is substantially reduced because of increased heterotrophic organism growth and associated oxygen consumption. The above loading condition is of concern mainly when primary effluent must be nitrified in conjunction with removing carbonaceous BOD.

Nitrification of secondary effluent, on the other hand, is governed mainly by the TKN loading to a Biocarbone unit. Between nitrogen loadings of 0.010 and 0.037 lb TKN/ft^3/day

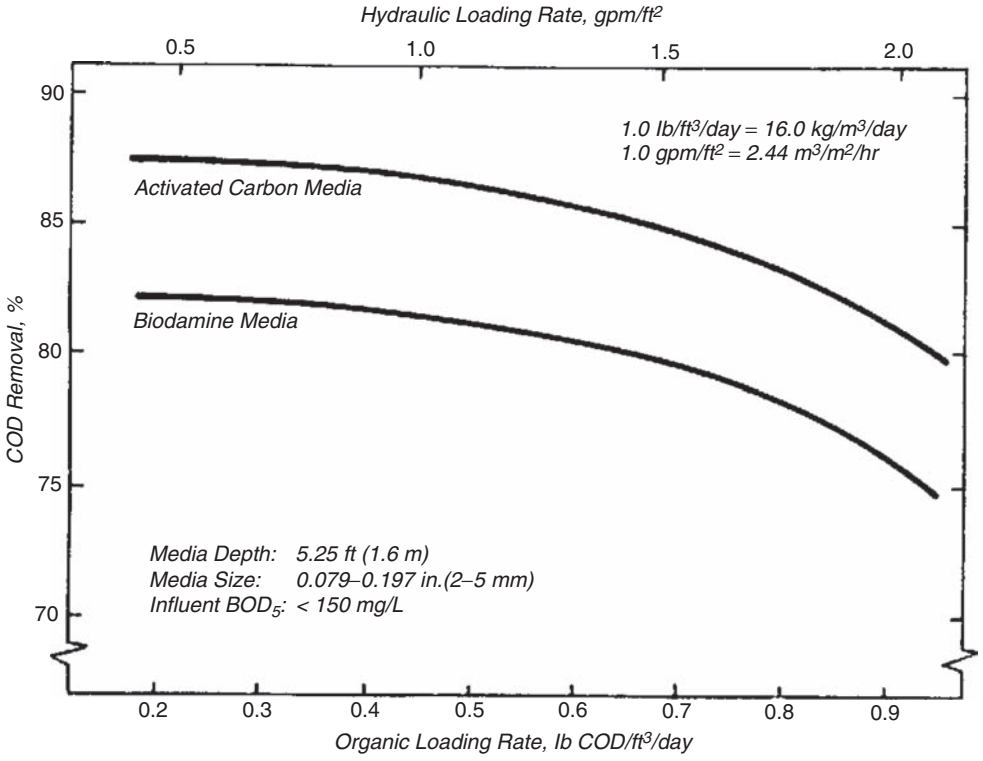


Fig. 17.10. COD removal as a function of BAF influent COD and hydraulic loading rates (50).

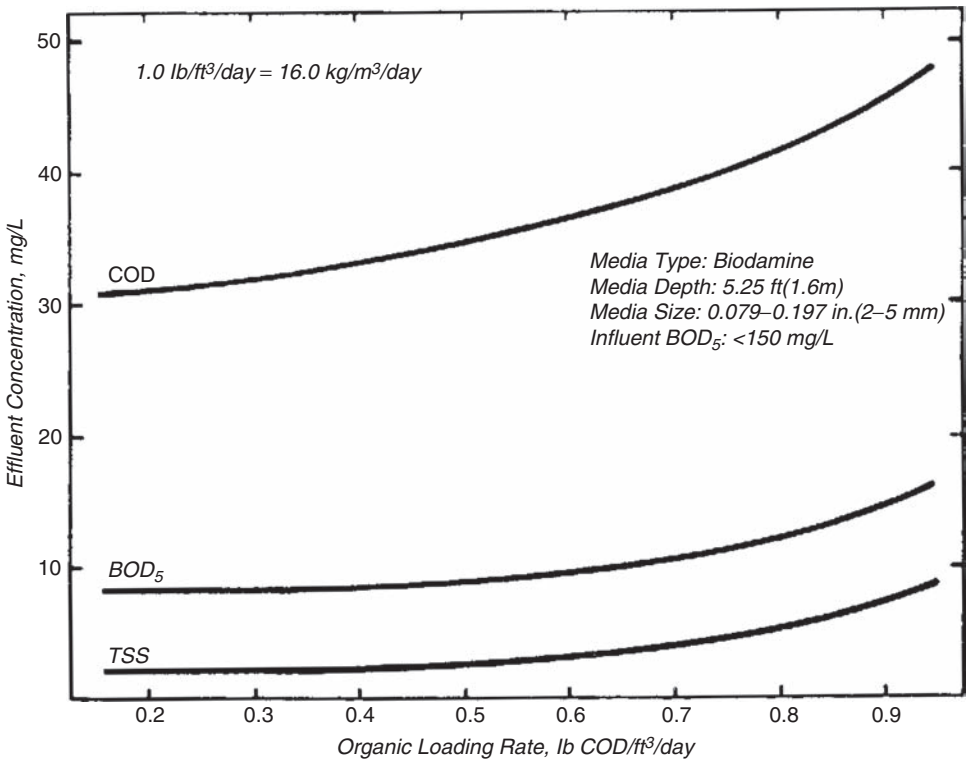


Fig. 17.11. BAF effluent quality as a function of influent COD loading rate (50).

(0.16 and 0.59 $kg/m^3/day$), NH-N removal decreases at a relatively linear rate, from about 90% to 84% (Fig. 17.12). Loadings above about 0.037 lb TKN/ft^3/day (0.59 $kg/m^3/day$) result in substantially reduced NH-N removal rates (50).

Based on data from another Biocarbone pilot plant study (52), a COD volumetric load of less than 2.0 $kg/m^3/day$ (0.125 $lb/ft^3/day$) was required to achieve approximately 90% ammonium oxidation in a single BAF unit. The BAF medium used in the pilot study was metamorphosed shale with a grain size between 3 and 6 mm (0.12 to 0.24 in.).

Pilot plant studies also provided data on the temperature dependence of NH-N oxidation. Based on ammonia-N oxidation in secondary effluent, OTV reported removal rates to approximate the following (50):

(a) At 12 °C (54 °F) the ammonia-N removal rate is 0.39 $kg/m^3/day$ (0.024 $lb/ft^3/day$)
(b) At 18 °C (64 °F) the ammonia-N removal rate is 0.50 $kg/m^3/day$ (0.031 $lb/ft^3/day$)
(c) At 24 °C (75 °F) the ammonia-N removal rate is 0.60 $kg/m^3/day$ (0.037 $lb/ft^3/day$)

According to results from the operation in the United States of a full-scale demonstration Biocarbone BAF plant treating primary municipal wastewater in the mid-1980s (53), the BOD_5 volumetric loading must be limited to approximately 1 $kg/m^3/day$ (0.0624 $lb/ft^3/day$) to achieve near 90% ammonium oxidation in a single unit. This conclusion is based on

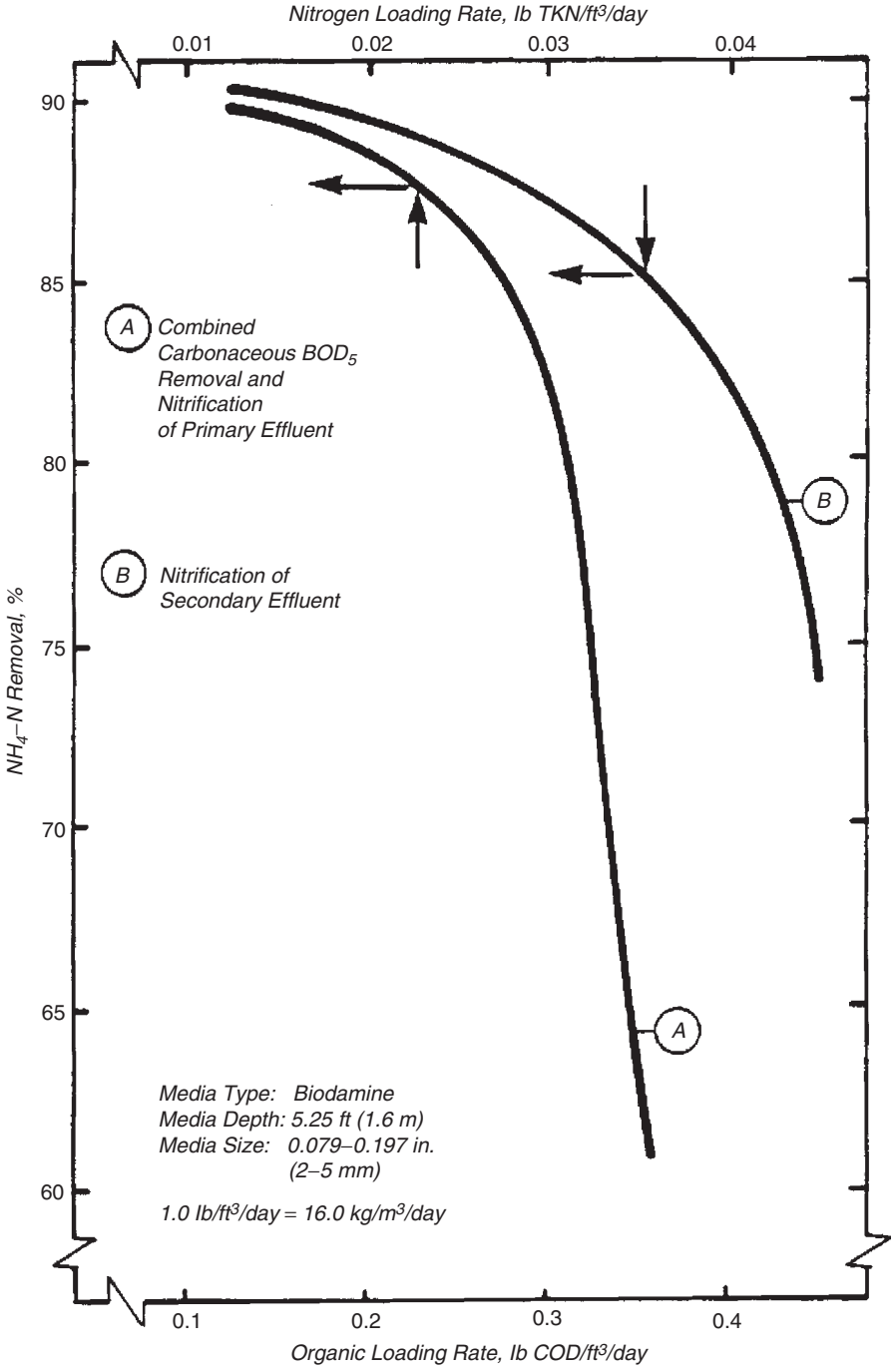


Fig. 17.12. Effect of COD loading on BAF nitrification performance (50).

Table 17.7
Design Parameters for Biological Aerated Filter (BAF)

Parameter	Units	Range
Organic loading	kg BOD/m^3/day	3–5
Hydraulic loading	m^3/m^2/day (m/day)	1–4
Contact time	h	0.5–1
Sludge production	kg SS/kg BOD	0.6–0.9
Bed height	M	2–3
Backwashing	m^3	2.5–3 × filter volume
Backwashing time	min daily	20
Energy consumption	kWh/kg BOD_5	1.0–1.3

operation at temperatures as low as 11 °C (52 °F) using a vitrified clay medium with an effective size of 3.4 mm (0.13 in.) and a uniformity coefficient between 1.5 and 1.6. Other more recent full-scale Biocarbone BAF plant assessments indicate that to achieve an average effluent ammonia-N concentration of 2.5 mg/L in the treatment of primary eff1uent, the COD volumetric loading must be limited to approximately 5 kg/m^3/day (0.312 lb/ft^3/day). The volumetric loading rate results indicate that carbonaceous oxidation and nearly complete nitrification of primary treated wastewater can be achieved in single BAF units at an empty-bed hydraulic retention time of approximately 1.5 to 3.5 h.

BAFs are typically designed to treat municipal wastewaters with low carbonaceous feed concentration, such as that characteristic of secondary effluent. In an EPA-sponsored, detailed assessment of BAFs (49), information derived from operation of a full-scale BAF unit treating secondary effluent was used to develop a design approach to predict the empty-bed hydraulic retention time required to achieve nitrification. At an influent BOD_5 and TSS concentration of approximately 20 mg/L, a hydraulic retention lime of 0.83 h was predicted to be required to reduce the ammonium nitrogen from approximately 21 to 7 mg/L. These results translate to an ammonium-nitrogen loading of 0.58 kg/m^3/day (0.036 lb/ft^3/day). Other reports indicate that over 90% removal of ammonium nitrogen is achievable at comparable volumetric loading rates at temperatures as tow as 13.5 °C (56.3 °F) (51). Design parameters extracted from various publications (50–63) are listed in Table 17.7.

Although full-scale application of BAFs for municipal wastewater treatment has become widespread in recent years, particularly in Europe (52), the amount of operating and performance information on US installations is limited. The lack at an extensive data base on nitrification applications suggests that onsite piloting may be warranted before selecting a technology (2).

3.5. Solids Production

The solids production rate in the BAF/Biocarbone process is a function of, among other factors, the quantities of soluble BOD, nonbiodegradable TSS, NH-N, and TKN removed. OTV initially used the historic solids production approximation of 0.7 to 0.8 lb solids/lb total

BOD_5 removed (kg/kg). A larger data base acquired from both pilot- and full- scale facilities yielded the following two modifications by OTV to their historic solids production value:

Solids production rate (50):

$= 0.4$ lb (kg)/lb (kg) soluble BOD_5 removed $+ 1.0$ lb (kg)/lb (kg) insoluble BOD_5 removed (1)
$= 0.4$ lb (kg)/lb (kg) soluble BOD_5 removed $+ 0.65$ lb (kg)/lb (kg) TSS removed (2)

Either of the above predicted models may be used to approximate the net solids production rate.

4. HYBRID BIOLOGICAL-ACTIVATED CARBON SYSTEMS

4.1. General Introduction

While the following processes were developed in laboratory experiments and verified in pilot studies in 1980s, they became popular only recently:

(a) First physicochemical fluidized bed GAC process.
(b) First biological fluidized bed GAC process.
(c) First physicochemical GAC sequencing batch reactor (SBR).
(d) First biological GAC-SBR.
(e) First combined dissolved air flotation (DAF) and GAC process.
(f) First DAF-PAC process.
(g) First physicochemical PAC-SBR process.
(h) First biological PAC-SBR process.
(i) First physicochemical PAC-DAF-SBR process.
(j) First biological PAC-DAF-SBR process.
(k) First ion exchange SBR process.
(l) First physicochemical SBR process.
(m) First regenerable gas phase GAC system.

Because of the importance of the above technologies, many US patents concerning GAC/PAC in combination with SBR, DAF and precoat filtration were filed by and granted to Wang and his co-workers (64–67).

The biological GAC filtration process was introduced as a competitive process to DAF-GAC process in 1989 (68). Mainstream Bio-Manipulation Systems Ltd., adapted both the slow sand filtration and biological GAC filtration processes in 1996 for drinking water production (69). In 2003, the first dual-stage biological GAC filtration plant is the 230-ML/day (230-million liters per day) Ngau Tam Mei Water Works, Hong Kong, China (70). In 2000, the first biological fluidized bed GAC system was built by both Envirogen and US Filter for groundwater decontamination (71).

4.2. Downflow Conventional Biological GAC Systems

4.2.1. Introduction

Granular activated carbon (GAC) adsorption system can remove many adsorble organics and inorganics, but not nonadsorble pollutants such as, dimethylnitrosamine, acetone cyanohydrin, butylamine, choline chloride, cyclohexylamine, diethylene glycol,

ethylenediamine, triethanolamine, and ethanol. Biological process, on the other hand, can remove biodegradable pollutants and not any nonbiodegradable pollutants. Combination of both processes will solve many traditionally unsolvable environmental pollution control problems.

It has been recognized by researchers and engineers that biological activity plays a major role in the removal of organics by activated carbon. When granular activated carbon is used simultaneously as the filtration and biological growth media in an attached growth biological oxidation-adsorption system, such a combination is called biological GAC adsorption system.

The conventional biological GAC process consists of a fixed bed of granular activated carbon media over which wastewater is applied for aerobic biological and adsorption treatment aiming at the removal of toxic organic substances. Biological slimes form on the GAC media, which assimilate and oxidize substances in the wastewater. The bed is dosed by a distributor system and the treated wastewater is collected by an underdrain system.

The organic material present in the wastewater is degraded by population of microorganisms attached to the GAC media and partially adsorbed by GAC macropores and micropores. The thickness of the slime layer increases as the microorganisms grow during the bio-oxidation process. The macropores and micropores of GAC are also gradually saturated by the target organic pollutants during adsorption. The microorganisms are also partially responsible for continuous GAC regeneration and prolonged adsorption. Periodically, the GAC bed must be backwashed and regenerated for reuse.

Both downflow pressurized biological GAC system and downflow gravity biological GAC system are technically feasible for water and wastewater treatment as long as oxygen is available for bio-oxidation (72, 73).

4.2.2. Saskatchewan-Canada Biological GAC Filtration Plant for Biological Treatment of Drinking Water

Slow sand filtration (water moves through such filters 10 to 20 times slower than in rapid sand filters) relies on the formation of a biological layer at the top of the filter. The filter does not become effective until this layer has been formed (68, 69). The American Water Works Association (AWWA) states: "The slow sand filtration process is expected to remove such biological particles as cysts, algae, bacteria, viruses, parasite eggs, nematode eggs, and amorphous organic debris at 100- to 10,000-fold levels when the filter is biologically mature". As effective as sand filtration can be, it is possible to maintain much greater numbers of microorganisms if the support material is GAC instead of sand. It is therefore preferable to use GAC for the removal of dissolved organics (69).

Mainstream Bio-Manipulation Systems Ltd., Canada, has, with the support of the National Research Council, worked on adapting both the slow sand filtration and biological GAC filtration processes. Such treatment systems have been installed at three different sites across Saskatchewan. One site has been in operation since 1996 and removal rates of turbidity, dissolved organic carbon, and color have been good for both the sand filter and the biological GAC filter. Both have provided high quality household water with no color or odor (removal rates of turbidity, dissolved organic carbon, and color are consistently above 50%). For drinking water purposes, the water is polished by a reverse osmosis unit. All of the household

water was hauled before installation of the biological treatment system. Based on successes like this one, it is anticipated that biological treatment will become one of the most common future treatment tools for dealing with surface waters on the Canadian prairie (69).

4.2.3. Ngau Tam Mei Water Works, Hong Kong, China

In 1994, facing projected shortfalls of potable water for the North Western New Territories of Hong Kong, the water supplies department initiated new facilities for treatment, conveyance and storage of water from its major supply, the Dongjiang River in Guangdong Province, People's Republic of China, via the Western Aqueduct.

In 2000, the Ngau Tam Mei water treatment works was commissioned, officially opening on December 2. It is the first water treatment plant worldwide to use dual-stage biological filtration with granular activated carbon (GAC) to remove ammonia, replacing break-point chlorination (70). The HKD 1.8 billion (USD 227 million) project treats river raw water, which is contaminated by wastewater. The plant was designed with an initial capacity of 230 ML/day, expandable to 450 ML/day.

The innovative plant was able to meet or surpass the required water quality goals by employing the following treatment units:

(a) Four pre-ozone contact tanks with a design detention time of 5 minutes.
(b) Twelve triple-deck sedimentation basins with a designed surface loading rate of 1.3 m/h.
(c) Intermediate ozone contact tanks with a design retention time of 15 min for achieving 1-log inactivation of Cryptospordium.
(d) Twelve first-stage GAC (1.5-m depth) filters with minimum filters run time of 24 h and filtration rate of 12 m/h, followed by 12 second-stage GAC (1.8-m depth) filters with a filtration rate of 8 m/h.
(e) Ozone peak dosage of 5 mg/L, ozone production rate of 1,150 kg/day and ozone concentration of 7.5%.

The plant was designed such that it is able to reduce O&M cost by:

(a) Generating high-quality oxygen on site, eliminating more costly truck-delivered liquid oxygen.
(b) Using dual-stage GAC filters to remove ammonia, eliminating break-point chlorination.
(c) Providing flexibility for operating in direct-filtration mode during periods of acceptable raw water quality to reduce coagulant chemical doses and sludge production.
(d) Reducing labor cost and improving plant management through a supervisory control and data acquisition (SCADA) system (70).

The three special advanced features of this largest biological GAC filtration plant include: (70)

(a) *Dual-stage biological GAC filtration*. A first-of-its-kind application in drinking water treatment. First-stage filters remove turbidity, biodegradable organic carbon, and taste- and odor-causing compounds. Second-stage filters remove ammonia, eliminating break-point chlorination and associated high chlorine doses. Results since commissioning show complete removal of ammonia (effluent concentration <0.02 mg/L).
(b) *Ozonation for primary disinfection*. This inactivates Giardia and Cryptospordium and reduces chlorine usage, helping to eliminate formation of chlorinated byproducts (THMs) and enhancing downstream biological filtration by oxygenating water and increasing formation of biodegradable organic carbon.

(c) *Ozonation for manganese removal*. Process uses preozonation for oxidation of reduced manganese to its insoluble form (manganese dioxide) for subsequent removal by coagulation and settling, followed by intermediate ozonation, which oxidizes remaining manganese in the settled water to permanganate for subsequent catalytic removal by first-stage GAC filters.

4.3. Upflow Fluidized Bed Biological GAC System (FBB-GAC)

Upflow fluidized bed biological GAC system has less clogging problem than the two downflow biological GAC systems introduced previously. Accordingly, the downflow biological GAC filtration process is mainly used for potable water treatment, whereas the upflow fluidized bed biological GAC system may be used for both water and wastewater treatment (68, 74). Many researchers are studying the upflow fluidized bed biological GAC systems (71, 74–76). The first fluidized bed-biological GAC system was designed and built in 2000 by Envirogen and US Filter for groundwater decontamination (71).

The FBB-GAC system (Hydroxyl Systems' Fluidized Bed Bioreactor) shown in Fig. 17.13 can be used in aerobic, anoxic or anaerobic conditions and can accommodate a variety of granular media (77). When adsorbent media such as granular activated carbon (GAC) is used, the FBB combines the benefits of adsorption and bio-oxidation. Contaminants are adsorbed onto the media surface and oxidized by the biofilm that is formed on the GAC surface. Unlike other biological treatment systems, the requirement for operator attention is minimal and unattended operation is practical. One of the most outstanding features of the FBB-GAC is that treatment detention times are typically minutes rather than hours.

The FBB-GAC system is supplied either as a single skid module of shippable height, incorporating a low profile reactor, or as a two-piece unit with a detachable tall cylindrical reactor. The system is used for aerobic, anoxic or anaerobic treatment of waterborne biodegradable matter, particularly adsorbable contaminants in low milligram per liter concentrations. Typical applications include treatment of groundwater contaminated with BTEX and as a

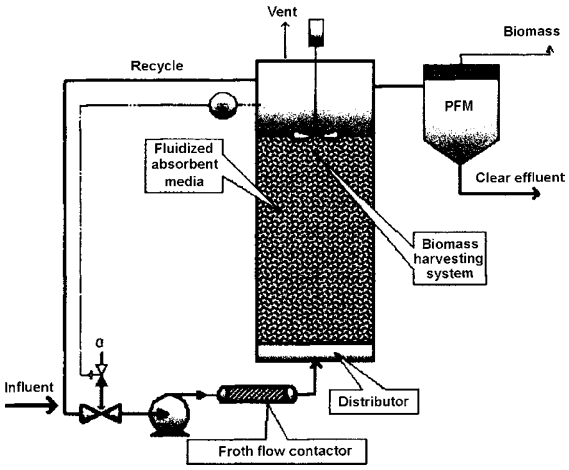


Fig. 17.13. Fluidized Bed Biological (FBB)-GAC System (77)

complement to advanced oxidation technologies for complete mineralization of biorefractory contaminants. As an anaerobic reactor, the FBB-GAC system can be used to treat high-strength wastewaters. Typical treated contaminants include BTEX, glycol, MTBE, soluble Oil & Grease and organic solvents.

The FBB-GAC system has the following special features (77):

(a) Fast bio-oxidation.
(b) Fully automated with PLC control.
(c) Weatherproof container (optional).
(d) No plugging or sludge bulking.
(e) No post-clarification required.
(f) Very compact and portable.
(g) Unattended operation.
(h) No off-gas.

An extremely high concentration of biomass develops in the reactor because of the huge surface area provided by the media, abundant oxygen and optimized mass transfer conditions. Excess biomass is periodically and automatically removed by extracting media, shearing the biomass and returning the cleaned biomass to the reactor. The effluent from the FBB-GAC is typically very low in suspended solids, allowing discharge without further treatment (77).

REFERENCES

1. Envirogen, Fluidized Bed Reactor- FBR, www.envirogen.com/fbrs.htm, (2004).
2. US EPA, *Nitrogen Control*, Tech. Report # EPA/625/R-93/010, U. S. Environmental Protection Agency, Washington, DC (1993).
3. P. M. Sutton and P. N. Mishra, Biological fluidized beds for water and wastewater treatment: a state of the art review, Presented at the *63rd Annual Conference of the Water Pollution Control Federation*, Washington, DC, October (1990).
4. N. J. Pugh, Treatment of doubtful waters for public supplies, *Transactions of Institution of Water Engineers* 50, 80 (1945).
5. US Patent No. 3,846,289 (1974).
6. W. J. Weber, Jr., et al., Physiochemical treatment of wastewater, *Journal of Water Pollution Control Federation*, 42, 83 (1970).
7. M. Rovatti, R. Di Felice, and C. Nicolella, Biomass concentration in fluidized bed biological reactor, *Water Research*, 31, 4, 936 (1987).
8. P. F. Cooper and B. Atkinson (eds.) *Biological Fluidized Bed Treatment of Water and Wastewater*, Ellis Horwood Publishers, Chichester, England (1981).
9. P. M. Sutton and P. N. Mishra, Biological fluidized beds for water and wastewater treatment, *Water Environment and Technology* 3, 8, 52 (1991).
10. S. S. Cheng, Organic carbon supplement influencing performance of biological nitrification in a fluidized bed reactor, *Water Science and Technology*, 30, 11, 131 (1994).
11. J. C. Araujo, J. R. Campos, and R. F. Vazoller, Methagonic biofilm: structure and microbial population activity in an anaerobic fluidized bed reactor treating synthetic wastewater, *Biofilm Journal*, 3, 1 (1998).
12. J. S. Jeris, R.W. Owens, R. Hickey, and F. Flood, Biological fluidized bed treatment for BOD and nitrogen removal, *Journal of Water Pollution Control Federation*, 49, 816 (1977).

13. Dearborn Environmental Consulting Services, *Pilot-Scale Assessment of the Biological Fluidized Bed for Municipal Wastewater Treatment*, November (1980).
14. P. M. Sutton, W. K. Shieh, P. Kos, and P. R. Dunning, Dorr-Oliver Oxitron System fluidized bed water and wastewater treatment process, In: *Biological Fluidized Bed Treatment of Water and Wastewater*, P. F. Cooper and B. Atkinson (eds.), Ellis Horwood Publishers, Chichester, England, p. 285 (1981).
15. D. K. Stevens, P. M. Berthouex, and T. W. Chapman, Dynamics and simulation of a biological fluidized bed reactor, *Proceedings of the First International Conference on Fixed-Film Biological Processes*, Kings Island, OH, 3, 1247 (1982).
16. M. K. Green and P. J. Hardy, The development of a high-rate nitrification fluidized bed process, *Water Pollution Control*, 84, 44 (1985).
17. P. F. Cooper and S. C. Williams, High-rate nitrification in a biological fluidized bed, *Water Science and Technology*, 22 (1/2), 431–442 (1990).
18. H. Tanaka and I. J. Dunn, Kinetics of biofilm nitrification, *Biotechnology and Bioengineering*, 34, 669 (1982).
19. W. K. Shieh, Predicting reactor biomass concentration in a fluidized bed system, *Journal of Water Pollution Control Federation*, 53, 1574 (1981).
20. J. S. Jeris and R. W. Owens, Pilot scale high rate biological denitrification, *Journal of Water Pollution Control Federation*, 47, 8 (1975).
21. N. K. Shammas, An allosteric kinetic model for the nitrification process, *Proceedings of the Tenth Annual Conference of Water Supply Improvement Association*, Honolulu, Hawaii, pp. 1–30, July (1982).
22. N. K. Shammas, Interactions of temperature, pH and biomass on the nitrification process, *Journal Water Pollution Control Federation*, 58, 1, 52–59, January (1986).
23. N. K. Shammas, An investigation of the effect of temperature on the oxidation of ammonia-n in municipal waste water treatment, *Fifth Annual Science Meeting*, Lebanese Association for the Advancement of Science, Beirut, Lebanon, December (1973).
24. N. K. Shammas, *Optimization of Biological Nitrification*, Ph.D. dissertation, Microfilm Publication, University of Michigan, Ann Arbor, MI (1971).
25. C. N. Sawyer, Activated sludge oxidations, V. The influence of nutrition in determining activated sludge characteristics, *Sewage Works Journal* 12, 3 (1940).
26. M. A. Moteleb, M. T. Suidan, J. Kim, J. L. Davel, and N. R. Adrian, Anaerobic degradation of 2,4,6-trinitrotoluene in a granular activated carbon fluidized bed and batch reactors, *Water Science and Technology*, 43, 1, 67 (2001).
27. U. S. Army Corps of Engineers, Fact Sheet: Pathways and Controlling Factors in the biodegradation of energetic wastes, construction engineering research laboratory, www.cecer.army.mil/facts/sheets/UL23.html, January (1998).
28. US Army Corps of Engineers, Fact Sheet: Anaerobic, *Granular-Activated Fluidized Bed Reactors for Treatment*, Engineering Research and Development center, www.cecer.army.mil (2004).
29. N. E. Davis (ed.), Combining new and established technologies means cleaner, safer water, *Centerpoint*, 4, 2, 8 (1998/99).
30. US Filter, *Groundwater Treatment: GAC Fluid Bed Systems*, http://hpi.usfilter.com/water_center (2003).
31. P. B. Hatzinger, et al., Biological treatment of perchlorate-contaminated groundwater using fluidized bed reactors, Presented at the *2nd International Conference on Remediation of Chlorinated and Recalcitrant Compounds*, Monterey, CA, May 22–25 (2000).

32. C. N Sawyer, H. E. Wild, Jr. and T. C. McMahon, *Nitrification and Denitrification Facilities, Wastewater Treatment*, Prepared for the EPA Technology Transfer Program (1973).
33. WPCF, *Nutrient Control*, Manual of Practice FD-7, Water Pollution Control Federation (1983).
34. WEF, *Design of Municipal Wastewater Treatment Plants*, Manual of Practice # 8, Water Environment Federation and American Society of Civil Engineers (1998).
35. Metcalf and Eddy, *Wastewater Engineering Treatment and Reuse*, 4th ed., McGraw Hill, New York (2003).
36. A. Vesilind, *Wastewater Treatment Plant Design*, Water Environment Federation and IWA Publishing, Alexandria, VA, USA (2003).
37. B. Ruston, Wastewater treatment with aerated submerged biological filters *Journal Water Pollution Control Federation*, 56, 424 (1984).
38. R. T. Haug and P. L. McCarty, Nitrification with the submerged filter *Journal of Water Pollution Control Federation*, 44, 2086 (1972).
39. U. S. EPA, Process Design Manual for Nitrogen Control, US Environmental Agency, Technology Transfer, October (1975).
40. M. Bonhomme, F. Rogalla, G. Boisseau, and J. Sibony, Enhancing nitrogen removal in activated sludge with fixed biomass, *Water Science and Technology*, 22(1/2), 127, 136 (1990).
41. S. Schlegel, The use of submerged biological filters for nitrification, *Water Science and Technology*, 20, 177 (1988).
42. O. Saintpierre, Tertiary nitrifying immerged biofilter with plastic media, *Environmental Technology Letters*, 9, 1059 (1988).
43. U. S. EPA, Innovative and Alternative Technology Assessment Manual, EPA430/9-78-009, US Environmental Protection Agency, Washington, DC (1980).
44. H. Furumai, H. Tagui, and K. Fujita, Effects of pH and alkalinity on sulfur-denitrification in a biological granular filter, *Water Science and Technology*, 34, 1–2, 355 (1996).
45. K. F. Janning, P. Harremoes, and M. Nielsen, Evaluating and modelling the kinetics in a full scale submerged denitrification filter, *Water Science and Technology*, 32, 8, 115 (1995).
46. R. A. Greene, Complete nitrification by single-stage activated sludge, Presented at the *46th Annual Conference of the Water Pollution Control Federation*, Cleveland, OH, October (1978).
47. US EPA, Energy Conservation in Municipal Wastewater Treatment, US Environmental Protection Agency, Report #. 430/9–77–011, March (1977).
48. US ACE, Civil Works Construction Cost Index System Manual, 110–2–1304, US Army Corps of Engineers, Washington, DC, PP 44, http://www.nww.usace.army.mil/cost. (2008).
49. A. J. Condren, *Technology Assessment of the Biological Aerated Filter*, US Environmental Protection Agency EPA/600/2-90/015, Cincinnati, OH, April (1990).
50. U. S. EPA, *Project Summary, Technology Assessment of the Biological Aerated Filter*, US Environmental Protection Agency EPA/600/S2-90/015, Cincinnati, OH, July (1990).
51. F. Rogalla and J. Sibony, Biocarbone aerated filters—ten years after: past, present, and plenty of potential, *Water Science and Technology*, 26, 9–11, 2043–2048 (1992).
52. F. Rogalla, M. Payraudeau, G. Bacquet, M. Bourbigot, J. Sibony, and P. Filles, Nitrification and phosphorous precipitation with biological aerated filters, *Journal Water Pollution Control Federation*, 62, 169 (1990).
53. H. D Stensel, R. C. Brenner, K. M. Lee, H. Melcer, and K. Rakness, Biological aerated filter evaluation, ASCE, *Journal of Environmental Engineering*, 3, 655 (1988).
54. F. Rogalla, M. Badard, and F. Hansen, Upscaling a compact nitrogen removal process, *Water Science Technology*, 26, 5–6, 1067–1076 (1992).

55. F. Rogalla and M. Bourbigot, New developments in complete nitrogen removal with biological aerated filters, *Water Science Technology*, 6, 1/2 (1990).
56. NE-WTTAC, GAC Biological Filtration, Technological Expertise of the New England Water Treatment Tecnology Assistance Center, www.unh.edu/erg/wttac/technological_expertise.htm March (2002).
57. G. Dillon and V. Thomas, A pilot-scale evaluation of the biocarbone process for the treatment of settled sewage and for tertiary nitrification of secondary effluent, *Water Science and Technology*, 22, 1/2, 305–316 (1990).
58. Gilles, S., Industrial-scale applications of fixed biomass on the Mediterranean seaboard, design, operating results, *Water Science and Technology*, 22, 1/2, 281–292 (1990).
59. K. Lee and D. Stensel, Aeration and substrate utilization in sparged packed bed biofilm reactor, *Journal Water Pollution Control Federation*, 58, 11, 1066–1072, November (1986).
60. B. Lacamp, F. Hansen, P. Penillard, and F. Rogalla, Wastewater nutrient removal with advanced biofilm reactors, *WEF, 65th Annual Conference*, New Orleans, 3, pp. 87–98, September (1992).
61. X. Zhu, M. T. Suidan, C. Alonso, T. Yu, B. J. Kim, and B. R. Kim, Biofilm structure and mass transfer in a gas phase trickle-bed biofilter, *Water Science and Technology*, 43, 1, 285 (2001).
62. B. Paffoni, B. Vedry, and M. Gousailles, Tertiary nitrification pilot plants on parisian wastewater, *Water Science and Technology*, 22, 1/2, 347–352 (1990).
63. R. Pujol, J. Canler, and A. Iwema, Biological aerated filters: an attractive and alternative biological process, *Water Science and Technology*, 26, 3–4, 693–702 (1992).
64. L. K. Wang, L. Kurylko, and M. H. Wang, *Improved Method and Apparatus for Liquid Treatment*, U. S. Patent and Trademark Office, Washington, DC, Patent No. 5256299, October 26 (1993).
65. L. K. Wang, L. Kurylko, and O. Hyrcyk, *Site Remediation Technology*, US Patent and Trademark Office, Washington, DC, Patent No. 5552051, Sept. (1996).
66. L. K. Wang and L. Kurylko, *Contamination Removal System Employing Filtration, Plural Ultraviolet & Chemical Treatment Steps & Treatment Controller*, US Patent No. 5,190,659; *Method and Apparatus for Filtration with Plural Ultraviolet Treatment Stages*, US Patent No. 5,236,595. US, Patent and Trademark Office, Washington, DC (1993).
67. L. K. Wang and L. Kurylko, *Sequencing Batch Liquid Treatment*, US Patent and Trademark Office, Washington, DC, US Patent No. 5,354,458, Oct. (1994).
68. L. K. Wang, Using air flotation and filtration in removal of color, trihalomethane precursors and Giardia Cysts, *NY State Department of Health Workshop on Water Treatment Chemicals and Filtration & the 1989 American Slow Sand Association Annual Meeting*, # P904–8–89–23, 30 p. (1989).
69. H. Peterson, Biological treatment of drinking water, *Critical Reviews in Environmental Control*, 2, 3, (1998) www.quantumlynx.com/water'back/vol5no2/story3.html.
70. P. Y. Chung, *Ngau Tam Mei Water Works, Hong Kong*, Camp Dresser & McKee, New York (2003).
71. US Filter, *Implementation of GAC Fluidized- Bed for Treatment of Petroleum Hydrocarbons in Groundwater at Two BP Oil Distribution Terminals, Pilot and Full-Scale*, US Filter, USA, www.usfilter.com (2000).
72. N. Davis, L. Erickson, and R. Hayter, Green cleanup technologies benefit communities, *Centerpoint*, Georgia Institute of Technology, GA., 4, 2 (1998).
73. O. Griffini, M. Bao, D. Burrini, D. Santianni, C. Barbieri, and F. Pantani, Removal of pesticides during the drinking water treatmen process at Florence water supply, Italy, *Journal Water SRT-Aqua*, 48, 177–185 (1999).

74. L. K. Wang, New dawn in development of adsorption technologies, *the 20th Annual Meeting of the Fine Particle Society Symposium on Activated Carbon Technology*, Boston, MA, Aug. (1989).
75. S. S. Cheng, Organic carbon supplement influencing performance of biological nitrification in fluidized bed reactor, *Water Science and Technology*, 30, 11, 131 (1994).
76. V. F. Medina, J. S. Devinny, and M. Ramaratnam, Biofiltration of toluene vapors in a carbon-medium biofilter, *Biological Unit Processes for Hazardous Waste Treatment, Third International In Situ and On-Site Bioremediation Symposium*, Battelle Press, Columbus, OH (1995).
77. Hydroxyl System, *Fluidized Biological Reactor*, Hydroxyl System Inc., 9800 McDonald Park Road, Sidney, BC, Canada, www.hydroxyl.com/products/biological/hydroxyl-fbb.html (2003).

APPENDIX

US yearly average cost index for utilities (48)

Year	Index	Year	Index
1967	100	1988	369.45
1968	104.83	1989	383.14
1969	112.17	1990	386.75
1970	119.75	1991	392.35
1971	131.73	1992	399.07
1972	141.94	1993	410.63
1973	149.36	1994	424.91
1974	170.45	1995	439.72
1975	190.49	1996	445.58
1976	202.61	1997	454.99
1977	215.84	1998	459.40
1978	235.78	1999	460.16
1979	257.20	2000	468.05
1980	277.60	2001	472.18
1981	302.25	2002	484.41
1982	320.13	2003	495.72
1983	330.82	2004	506.13
1984	341.06	2005	516.75
1985	346.12	2006	528.12
1986	347.33	2007	539.74
1987	353.35	2008	552.16

Appendix: Conversion Factors for Environmental Engineers

Lawrence K. Wang

CONTENTS

Abstract With the current trend toward metrication, the question of using a consistent system of units has been a problem. Wherever possible, the authors of this *Handbook of Environmental Engineering* series have used the British system (fps) along with the metric equivalent (mks, cgs, or SIU) or vice versa. For the convenience of the readers around the world, this book provides a detailed Conversion Factors for Environmental Engineers. In addition, the basic and supplementary units, the derived units and quantities, important physical constants, the properties of water, and the Periodic Table of the elements, are also presented in this document.

Key Words Conversion factors • British units • metric units • physical constants • water properties • periodic table of the elements • environmental engineers • Lenox Institute of Water Technology • mks (meter-kilogram-second) • cgs (centimeter-gram-second) • SIU (Système international d'unités; International System of Units) • fps (foot-pound-second).

From: *Handbook of Environmental Engineering, Volume 9: Advanced Biological Treatment Processes*
Edited by: L. K. Wang, N. K. Shammas and Y-T. Hung, DOI: 10.1007/978-1-60327-170-7_BM © Humana Press, New York, NY 2009

1. CONSTANTS AND CONVERSION FACTORS

Multiply	*by*	*to obtain*
abamperes	10	amperes
abamperes	2.99796×10^{10}	statamperes
abampere-turns	12.566	gilberts
abcoulombs	10	coulombs (abs)
abcoulombs	2.99796×10^{10}	statcoulombs
abcoulombs/kg	30,577	statcoulombs/dyne
abfarads	1×10^{9}	farads (abs)
abfarads	8.98776×10^{20}	statfarads
abhenries	1×10^{-9}	henries (abs)
abhenries	1.11263×10^{-21}	stathenries
abohms	1×10^{-9}	ohms (abs)
abohms	1.11263×10^{-21}	statohms
abvolts	3.33560×10^{-11}	statvolts
abvolts	1×10^{-8}	volts (abs)
abvolts/centimeters	2.540005×10^{-8}	volts (abs)/inch
acres	0.4046	ha
acres	43,560	square feet
acres	4047	square meters
acres	1.562×10^{-3}	square miles
acres	4840	square yards
acre-feet	43,560	cubic feet
acre-feet	1233.5	cubic meters
acre-feet	325,850	gallons (U.S.)
amperes (abs)	0.1	abamperes
amperes (abs)	1.036×10^{-5}	faradays/second
amperes (abs)	2.9980×10^{9}	statamperes
ampere-hours (abs)	3600	coulombs (abs)
ampere-hours	0.03731	faradays
amperes/sq cm	6.452	amps/sq in
amperes/sq cm	10^{4}	amps/sq meter
amperes/sq in	0.1550	amps/sq cm
amperes/sq in	1550.0	amps/sq meter
amperes/sq meter	10^{-4}	amps/sq cm
amperes/sq meter	6.452×10^{-4}	amps/sq in
ampere-turns	1.257	gilberts
ampere-turns/cm	2.540	amp-turns/in
ampere-turns/cm	100.0	amp-turns/meter
ampere-turns/cm	1.257	gilberts/cm
ampere-turns/in	0.3937	amp-turns/cm
ampere-turns/in	39.37	amp-turns/meter
ampere-turns/in	0.4950	gilberts/cm

Multiply	*by*	*to obtain*
ampere-turns/meter	0.01	amp-turns/cm
ampere-turns/meter	0.0254	amp-turns/in
ampere-turns/meter	0.01257	gilberts/cm
angstrom units	1×10^{-8}	centimeters
angstrom units	3.937×10^{-9}	inches
angstrom unit	1×10^{-10}	meter
angstrom unit	1×10^{-4}	micron or μm
ares	0.02471	acre (U.S.)
ares	1076	square feet
ares	100	square meters
ares	119.60	sq yards
assay tons	29.17	grams
astronomical unit	1.495×10^{8}	kilometers
atmospheres (atm)	0.007348	tons/sq inch
atmospheres	76.0	cms of mercury
atmospheres	1.01325×10^{6}	dynes/square centimeter
atmospheres	33.90	ft of water (at 4°C)
atmospheres	29.92	inches of mercury (at 0°C)
atmospheres	1.033228	kg/sq cm
atmospheres	10,332	kg/sq meter
atmospheres	760.0	millimeters of mercury
atmospheres	14.696	pounds/square inch
atmospheres	1.058	tons/sq foot
avograms	1.66036×10^{-24}	grams
bags, cement	94	pounds of cement
barleycorns (British)	1/3	inches
barleycorns (British)	8.467×10^{-3}	meters
barrels (British, dry)	5.780	cubic feet
barrels (British, dry)	0.1637	cubic meters
barrels (British, dry)	36	gallons (British)
barrels, cement	170.6	kilograms
barrels, cement	376	pounds of cement
barrels, cranberry	3.371	cubic feet
barrels, cranberry	0.09547	cubic meters
barrels, oil	5.615	cubic feet
barrels, oil	0.1590	cubic meters
barrels, oil	42	gallons (U.S.)
barrels, (U.S., dry)	4.083	cubic feet
barrels (U.S., dry)	7056	cubic inches
barrels (U.S., dry)	0.11562	cubic meters
barrels (U.S., dry)	105.0	quarts (dry)
barrels (U.S., liquid)	4.211	cubic feet
barrels (U.S., liquid)	0.1192	cubic meters
barrels (U.S., liquid)	31.5	gallons (U.S.)

Multiply	*by*	*to obtain*
bars	0.98692	atmospheres
bars	10^6	dynes/sq cm
bars	1.0197×10^4	kg/sq meter
bars	1000	millibar
bars	750.06	mm of Hg (0°C)
bars	2089	pounds/sq ft
bars	14.504	pounds/sq in
barye	1.000	dynes/sq cm
board feet	1/12	cubic feet
board feet	144 sq.in. × 1 in.	cubic inches
boiler horsepower	33,475	BTU (mean)/hour
boiler horsepower	34.5	pounds of water evaporated from and at 212°F (per hour)
bolts (U.S., cloth)	120	linear feet
bolts (U.S., cloth)	36.576	meters
bougie decimales	1	candles (int)
BTU (mean)	251.98	calories, gram (g. cal)
BTU (mean)	0.55556	centigrade heat units (chu)
BTU (mean)	1.0548×10^{10}	ergs
BTU (mean)	777.98	foot-pounds
BTU (mean)	3.931×10^{-4}	horsepower-hrs (hp-hr)
BTU (mean)	1055	joules (abs)
BTU (mean)	0.25198	kilograms, cal (kg cal)
BTU (mean)	107.565	kilogram-meters
BTU (mean)	2.928×10^{-4}	kilowatt-hr (Kwh)
BTU (mean)	10.409	liter-atm
BTU (mean)	6.876×10^{-5}	pounds of carbon to CO_2
BTU (mean)	0.29305	watt-hours
BTU (mean)/cu ft	37.30	joule/liter
BTU/hour	0.2162	foot-pound/sec
BTU/hour	0.0700	gram-cal/sec
BTU/hour	3.929×10^{-4}	horsepower-hours (hp-hr)
BTU/hour	0.2930711	watt (w)
BTU/hour (feet)°F	1.730735	joule/sec (m)°k
BTU/hour (feet^2)	3.15459	joule/m^2-sec
BTU (mean)/hour(feet^2)°F	1.3562×10^{-4}	gram-calorie/second (cm^2)°C
BTU (mean)/hour(feet^2)°F	3.94×10^{-4}	horsepower/(ft^2)°F
BTU (mean)/hour(feet^2)°F	5.678264	joule/sec (m^2)°k
BTU (mean)/hour(feet^2)°F	4.882	kilogram-calorie/hr (m^2)°C
BTU (mean)/hour(feet^2)°F	5.682×10^{-4}	watts/(cm^2)°C
BTU (mean)/hour(feet^2)°F	2.035×10^{-3}	watts/(in^2)°C
BTU (mean)/(hour)(feet^2) (°F/inch)	3.4448×10^{-4}	calories, gram (15°C)/sec (cm^2) (°C/cm)
BTU (mean)/(hour)(feet^2) (°F/in.)	1	chu/(hr)(ft^2)(°C/in)

Multiply	*by*	*to obtain*
BTU (mean)/(hour)(feet^2) (°F/inch)	1.442×10^{-3}	joules (abs)/(sec)(cm^2) (°C/cm)
BTU (mean)/(hour)(feet^2) (°F/inch)	1.442×10^{-3}	watts/(cm^2) (°C/cm)
BTU/min	12.96	ft lb/sec
BTU/min	0.02356	hp
BTU/min	0.01757	kw
BTU/min	17.57	watts
BTU/min/ft^2	0.1221	watts/sq inch
BTU/pound	0.5556	calories-gram(mean)/gram
BTU/pound	0.555	kg-cal/kg
BTU/pound/°F	1	calories, gram/gram/°C
BTU/pound/°F	4186.8	joule/kg/°k
BTU/second	1054.350	watt (W)
buckets (British, dry)	1.818×10^4	cubic cm
buckets (British, dry)	4	gallons (British)
bushels (British)	1.03205	bushels (U.S.)
bushels (British)	1.2843	cubic feet
bushels (British)	0.03637	cubic meters
bushels (U.S.)	1.2444	cubic feet
bushels (U.S.)	2150.4	cubic inch
bushels (U.S.)	0.035239	cubic meters
bushels (U.S.)	35.24	liters (L)
bushels (U.S.)	4	pecks (U.S.)
bushels (U.S.)	64	pints (dry)
bushels (U.S.)	32	quarts (dry)
butts (British)	20.2285	cubic feet
butts (British)	126	gallons (British)
cable lengths	720	feet
cable lengths	219.46	meters
calories (thermochemical)	0.999346	calories (Int. Steam Tables)
calories, gram (g. cal or simply cal.)	3.9685×10^{-3}	BTU (mean)
calories, gram (mean)	0.001459	cubic feet atmospheres
calories, gram (mean)	4.186×10^7	ergs
calories, gram (mean)	3.0874	foot-pounds
calories, gram (mean)	4.186	joules (abs)
calories, gram (mean)	0.001	kg cal (calories, kilogram)
calories, gram (mean)	0.42685	kilograms-meters
calories, gram (mean)	0.0011628	watt-hours
calories, gram (mean)/gram	1.8	BTU (mean)/pound
cal/gram-°C	4186.8	joule/kg-°k
candle power (spherical)	12.566	lumens
candles (int)	0.104	carcel units
candles (int)	1.11	hefner units
candles (int)	1	lumens (int)/steradian
candles (int)/square centimeter	2919	foot-lamberts

Multiply	*by*	*to obtain*
candles (int)/square centimeter	3.1416	lamberts
candles (int)/square foot	3.1416	foot-lamberts
candles (int)/square foot	3.382×10^{-3}	lamberts
candles (int)/square inch	452.4	foot-lamberts
candles (int)/square inch	0.4870	lamberts
candles (int)/square inch	0.155	stilb
carats (metric)	3.0865	grains
carats (metric)	0.2	grams
centals	100	pounds
centares (centiares)	1.0	sq meters
centigrade heat units (chu)	1.8	BTU
centigrade heat units (chu)	453.6	calories, gram (15°C)
centigrade heat units (chu)	1897.8	joules (abs)
centigrams	0.01	grams
centiliters	0.01	liters
centimeters	0.0328083	feet (U.S.)
centimeters	0.3937	inches (U.S.)
centimeters	0.01	meters
centimeters	6.214×10^{-6}	miles
centimeters	10	millimeters
centimeters	393.7	mils
centimeters	0.01094	yards
cm of mercury	0.01316	atm
cm of mercury	0.4461	ft of water
cm of mercury	136.0	kg/square meter
cm of mercury	1333.22	newton/meter^2 (N/m^2)
cm of mercury	27.85	psf
cm of mercury	0.1934	psi
cm of water (4°C)	98.0638	newton/meter^2 (N/m^2)
centimeters-dynes	1.020×10^{-3}	centimeter-grams
centimeter-dynes	1.020×10^{-8}	meter-kilograms
centimeter-dynes	7.376×10^{-8}	pound-feet
centimeter-grams	980.7	centimeter-dynes
centimeter-grams	10^{-5}	meter-kilograms
centimeter-grams	7.233×10^{-5}	pound-feet
centimeters/second	1.969	fpm (ft/min)
centimeters/second	0.0328	fps (ft/sec)
centimeters/second	0.036	kilometers/hour
centimeters/second	0.1943	knots
centimeters/second	0.6	m/min
centimeters/second	0.02237	miles/hour
centimeters/second	3.728×10^{-4}	miles/minute
cms/sec./sec.	0.03281	feet/sec/sec
cms/sec./sec.	0.036	kms/hour/sec

Multiply	*by*	*to obtain*
cms/sec./sec.	0.02237	miles/hour/sec
centipoises	3.60	kilograms/meter hour
centipoises	10^{-3}	kilograms/meter second
centipoises	0.001	newton-sec/m^2
centipoises	2.089×10^{-5}	pound force second/square foot
centipoises	2.42	pounds/foot hour
centipoises	6.72×10^{-4}	pounds/foot second
centistoke	1.0×10^{-6}	$meter^2/sec$
chains (engineers' or Ramden's)	100	feet
chains (engineers' or Ramden's)	30.48	meters
chains (surveyors' or Gunter's)	66	feet
chains (surveyors' or Gunter's)	20.12	meters
chaldrons (British)	32	bushels (British)
chaldrons (U.S.)	36	bushels (U.S.)
cheval-vapours	0.9863	horsepower
cheval-vapours	735.5	watts (abs)
cheval-vapours heures	2.648×10^6	joules (abs)
chu/(hr)(ft^2)(°C/in.)	1	BTU/(hr)(ft^2)(°F/in.)
circular inches	0.7854	square inches
circular millimeters	7.854×10^{-7}	square meters
circular mils	5.067×10^{-6}	square centimeters
circular mils	7.854×10^{-7}	square inches
circular mils	0.7854	square mils
circumferences	360	degrees
circumferences	400	grades
circumferences	6.283	radians
cloves	8	pounds
coombs (British)	4	bushels (British)
cords	8	cord feet
cords	$8' \times 4' \times 4'$	cubic feet
cords	128	cubic feet
cords	3.625	cubic meters
cord-feet	$4' \times 4' \times 1'$	cubic feet
coulombs (abs)	0.1	abcoulombs
coulombs (abs)	6.281×10^{18}	electronic charges
coulombs (abs)	2.998×10^9	statcoulombs
coulombs (abs)	1.036×10^{-5}	faradays
coulombs/sq cm	64.52	coulombs/sq in
coulombs/sq cm	10^4	coulombs/sq meter
coulombs/sq in	0.1550	coulombs/sq cm
coulombs/sq in	1550	coulombs/sq meter
coulombs/sq meter	10^{-4}	coulombs/sq cm
coulombs/sq meter	6.452×10^{-4}	coulombs/sq in
cubic centimeters	3.531445×10^{-5}	cubic feet (U.S.)

Multiply	*by*	*to obtain*
cubic centimeters	6.102×10^{-2}	cubic inches
cubic centimeters	10^{-6}	cubic meters
cubic centimeters	1.308×10^{-6}	cubic yards
cubic centimeters	2.6417×10^{-4}	gallons (U.S.)
cubic centimeters	0.001	liters
cubic centimeters	0.033814	ounces (U.S., fluid)
cubic centimeters	2.113×10^{-3}	pints (liq.)
cubic centimeters	1.057×10^{-3}	quarts (liq.)
cubic feet (British)	0.9999916	cubic feet (U.S.)
cubic feet (U.S.)	0.8036	bushels (dry)
cubic feet (U.S.)	28317.016	cubic centimeters
cubic feet (U.S.)	1728	cubic inches
cubic feet (U.S.)	0.02832	cubic meters
cubic feet (U.S.)	0.0370	cubic yard
cubic feet (U.S.)	7.48052	gallons (U.S.)
cubic feet (U.S.)	28.31625	liters
cubic feet (U.S.)	59.84	pints (liq.)
cubic feet (U.S.)	29.92	quarts (liq.)
cubic feet of common brick	120	pounds
cubic feet of water (60°F)	62.37	pounds
cubic foot-atmospheres	2.7203	BTU (mean)
cubic foot-atmospheres	680.74	calories, gram (mean)
cubic foot-atmospheres	2116	foot-pounds
cubic foot-atmospheres	2869	joules (abs)
cubic foot-atmospheres	292.6	kilogram-meters
cubic foot-atmospheres	7.968×10^{-4}	kilowatt-hours
cubic feet/hr	0.02832	m^3/hr
cubic feet/minute	472.0	cubic cm/sec
cubic feet/minute	1.6992	cu m/hr
cubic feet/minute	0.0283	cu m/min
cubic feet/minute	0.1247	gallons/sec
cubic feet/minute	0.472	liter/sec
cubic feet/minute	62.4	lbs of water/min
cubic feet/min/1000 cu ft	0.01667	liter/sec/cu m
cubic feet/second	1.9834	acre-feet/day
cubic feet/second	1.7	cu m/min
cubic feet/second	0.02832	m^3/sec
cubic feet/second	448.83	gallons/minute
cubic feet/second	1699	liter/min
cubic feet/second	28.32	liters/sec
cubic feet/second (cfs)	0.64632	million gallons/day (MGD)
cfs/acre	0.07	m^3/sec-ha
cfs/acre	4.2	cu m/min/ha
cfs/sq mile	0.657	cu m/min/sq km

Multiply	*by*	*to obtain*
cubic inches (U.S.)	16.387162	cubic centimeters
cubic inches (U.S.)	5.787×10^{-4}	cubic feet
cubic inches (U.S.)	1.0000084	cubic inches (British)
cubic inches (U.S.)	1.639×10^{-5}	cubic meters
cubic inches (U.S.)	2.143×10^{-5}	cubic yards
cubic inches (U.S.)	4.329×10^{-3}	gallons (U.S.)
cubic inches (U.S.)	1.639×10^{-2}	liters
cubic inches (U.S.)	16.39	mL
cubic inches (U.S.)	0.55411	ounces (U.S., fluid)
cubic inches (U.S.)	0.03463	pints (liq.)
cubic inches (U.S.)	0.01732	quarts (liq.)
cubic meters	8.1074×10^{-4}	acre-feet
cubic meters	8.387	barrels (U.S., liquid)
cubic meters	28.38	bushels (dry)
cubic meters	10^6	cubic centimeters
cubic meters	35.314	cubic feet (U.S.)
cubic meters	61,023	cubic inches (U.S.)
cubic meters	1.308	cubic yards (U.S.)
cubic meters	264.17	gallons (U.S.)
cubic meters	1000	liters
cubic meters	2113	pints (liq.)
cubic meters (m^3)	1057	quarts (liq.)
cubic meters/day	0.183	gallons/min
cubic meters/ha	106.9	gallons/acre
cubic meters/hour	0.2272	gallons/minute
cubic meters/meter-day	80.53	gpd/ft
cubic meters/minute	35.314	cubic ft/minute
cubic meters/second	35.314	cubic ft/sec
cubic meters/second	22.82	MGD
cubic meters/sec-ha	14.29	cu ft/sec-acre
cubic meters/$meters^2$-day	24.54	gpd/ft^2
cubic yards (British)	0.9999916	cubic yards (U.S.)
cubic yards (British)	0.76455	cubic meters
cubic yards (U.S.)	7.646×10^5	cubic centimeters
cubic yards (U.S.)	27	cubic feet (U.S.)
cubic yards (U.S.)	46,656	cubic inches
cubic yards (U.S.)	0.76456	cubic meters
cubic yards (U.S.)	202.0	gallons (U.S.)
cubic yards (U.S.)	764.6	liters
cubic yards (U.S.)	1616	pints (liq.)
cubic yards (U.S.)	807.9	quarts (liq.)
cubic yards of sand	2700	pounds
cubic yards/minute	0.45	cubic feet/second
cubic yards/minute	3.367	gallons/second

Multiply	*by*	*to obtain*
cubic yards/minute	12.74	liters/second
cubits	45.720	centimeters
cubits	1.5	feet
dalton	1.65×10^{-24}	gram
days	1440	minutes
days	86,400	seconds
days (sidereal)	86164	seconds (mean solar)
debye units (dipole moment)	10^{18}	electrostatic units
decigrams	0.1	grams
deciliters	0.1	liters
decimeters	0.1	meters
degrees (angle)	60	minutes
degrees (angle)	0.01111	quadrants
degrees (angle)	0.01745	radians
degrees (angle)	3600	seconds
degrees/second	0.01745	radians/seconds
degrees/second	0.1667	revolutions/min
degrees/second	0.002778	revoltuions/sec
degree Celsius	$°F = (°C \times 9/5) + 32$	Fahrenheit
degree Celsius	$°K = °C + 273.15$	Kelvin
degree Fahrenheit	$°C = (°F - 32) \times 5/9$	Celsius
degree Fahrenheit	$°K = (°F + 459.67)/1.8$	Kelvin
degree Rankine	$°K = °R/1.8$	Kelvin
dekagrams	10	grams
dekaliters	10	liters
dekameters	10	meters
drachms (British, fluid)	3.5516×10^{-6}	cubic meters
drachms (British, fluid)	0.125	ounces (British, fluid)
drams (apothecaries' or troy)	0.1371429	ounces (avoirdupois)
drams (apothecaries' or troy)	0.125	ounces (troy)
drams (U.S., fluid or apoth.)	3.6967	cubic cm
drams (avoirdupois)	1.771845	grams
drams (avoirdupois)	27.3437	grains
drams (avoirdupois)	0.0625	ounces
drams (avoirdupois)	0.00390625	pounds (avoirdupois)
drams (troy)	2.1943	drams (avoirdupois)
drams (troy)	60	grains
drams (troy)	3.8879351	grams
drams (troy)	0.125	ounces (troy)
drams (U.S., fluid)	3.6967×10^{-6}	cubic meters
drams (U.S., fluid)	0.125	ounces (fluid)
dynes	0.00101972	grams

Multiply	*by*	*to obtain*
dynes	10^{-7}	joules/cm
dynes	10^{-5}	joules/meter (newtons)
dynes	1.020×10^{-6}	kilograms
dynes	1×10^{-5}	newton (N)
dynes	7.233×10^{-5}	poundals
dynes	2.24809×10^{-6}	pounds
dyne-centimeters (torque)	7.3756×10^{-8}	pound-feet
dynes/centimeter	1	ergs/square centimeter
dynes/centimeter	0.01	ergs/square millimeter
dynes/square centimeter	9.8692×10^{-7}	atmospheres
dynes/square centimeter	10^{-6}	bars
dynes/square centimeter	2.953×10^{-5}	inch of mercury at 0°C
dynes/square centimeter	4.015×10^{-4}	inch of water at 4°C
dynes/square centimeter	0.01020	kilograms/square meter
dynes/square centimeter	0.1	newtons/square meter
dynes/square centimeter	1.450×10^{-5}	pounds/square inch
electromagnetic fps units of magnetic permeability	0.0010764	electromagnetic cgs units of magnetic permeability
electromagnetic fps units of magnetic permeability	1.03382×10^{-18}	electrostatic cgs units of magnetic permeability
electromagnetic cgs units, of magnetic permeability	1.1128×10^{-21}	electrostatic cgs units of magnetic permeability
electromagnetic cgs units of mass resistance	9.9948×10^{-6}	ohms (int)-meter-gram
electronic charges	1.5921×10^{-19}	coulombs (abs)
electron-volts	1.6020×10^{-12}	ergs
electron-volts	1.0737×10^{-9}	mass units
electron-volts	0.07386	rydberg units of energy
electronstatic cgs units of Hall effect	2.6962×10^{31}	electromagnetic cgs units of Hall effect
electrostatic fps units of charge	1.1952×10^{-6}	coulombs (abs)
electrostatic fps units of magnetic permeability	929.03	electrostatic cgs units of magnetic permeability
ells	114.30	centimeters
ells	45	inches
ems, pica (printing)	0.42333	centimeters
ems, pica (printing)	1/6	inches
ergs	9.4805×10^{-11}	BTU (mean)
ergs	2.3889×10^{-8}	calories, gram (mean)
ergs	1	dyne-centimeters
ergs	7.3756×10^{-8}	foot-pounds
ergs	0.2389×10^{-7}	gram-calories
ergs	1.020×10^{-3}	gram-centimeters

Multiply	*by*	*to obtain*
ergs	3.7250×10^{-14}	horsepower-hrs
ergs	10^{-7}	joules (abs)
ergs	2.390×10^{-11}	kilogram-calories (kg cal)
ergs	1.01972×10^{-8}	kilogram-meters
ergs	0.2778×10^{-13}	kilowatt-hrs
ergs	0.2778×10^{-10}	watt-hours
ergs/second	5.692×10^{-9}	BTU/min
ergs/second	4.426×10^{-6}	foot-pounds/min
ergs/second	7.376×10^{-8}	foot-pounds/sec
ergs/second	1.341×10^{-10}	horsepower
ergs/second	1.434×10^{-9}	kg-calories/min
ergs/second	10^{-10}	kilowatts
farad (international of 1948)	0.9995	farad (F)
faradays	26.80	ampere-hours
faradays	96,500	coulombs (abs)
faradays/second	96,500	amperes (abs)
farads (abs)	10^{-9}	abfarads
farads (abs)	10^{6}	microfarads
farads (abs)	8.9877×10^{11}	statfarads
fathoms	6	feet
fathom	1.829	meter
feet (U.S.)	1.0000028	feet (British)
feet (U.S.)	30.4801	centimeters
feet (U.S.)	12	inches
feet (U.S.)	3.048×10^{-4}	kilometers
feet (U.S.)	0.30480	meters
feet (U.S.)	1.645×10^{-4}	miles (naut.)
feet (U.S.)	1.893939×10^{-4}	miles (statute)
feet (U.S.)	304.8	millimeters
feet (U.S.)	1.2×10^{4}	mils
feet (U.S.)	1/3	yards
feet of air (1 atmosphere, 60°F)	5.30×10^{-4}	pounds/square inch
feet of water	0.02950	atm
feet of water	0.8826	inches of mercury
feet of water at 39.2°F	0.030479	kilograms/square centimeter
feet of water at 39.2°F	2988.98	newton/meter^2 (N/m^2)
feet of water at 39.2°F	304.79	kilograms/square meter
feet of water	62.43	pounds/square feet (psf)
feet of water at 39.2°F	0.43352	pounds/square inch (psi)
feet/hour	0.08467	mm/sec
feet/min	0.5080	cms/sec
feet/min	0.01667	feet/sec
feet/min	0.01829	km/hr
feet/min	0.3048	meters/min

Multiply	*by*	*to obtain*
feet/min	0.01136	miles/hr
feet/sec	30.48	cm/sec
feet/sec	1.097	km/hr
feet/sec	0.5921	knots
feet/sec	18.29	meters/min
feet/sec	0.6818	miles/hr
feet/sec	0.01136	miles/min
feet/sec/sec	30.48	cm/sec/sec
feet/sec/sec	1.097	km/hr/sec
feet/sec/sec	0.3048	meters/sec/sec
feet/sec/sec	0.6818	miles/hr/sec
feet/100 feet	1.0	percent grade
firkins (British)	9	gallons (British)
firkins (U.S.)	9	gallons (U.S.)
foot-candle (ft-c)	10.764	lumen/sq m
foot-poundals	3.9951×10^{-5}	BTU (mean)
foot-poundals	0.0421420	joules (abs)
foot-pounds	0.0012854	BTU (mean)
foot-pounds	0.32389	calories, gram (mean)
foot-pounds	1.13558×10^{7}	ergs
foot-pounds	32.174	foot-poundals
foot-pounds	5.050×10^{-7}	hp-hr
foot-pounds	1.35582	joules (abs)
foot-pounds	3.241×10^{-4}	kilogram-calories
foot-pounds	0.138255	kilogram-meters
foot-pounds	3.766×10^{-7}	kwh
foot-pounds	0.013381	liter-atmospheres
foot-pounds	3.7662×10^{-4}	watt-hours (abs)
foot-pounds/minute	1.286×10^{-3}	BTU/minute
foot-pounds/minute	0.01667	foot-pounds/sec
foot-pounds/minute	3.030×10^{-5}	hp
foot-pounds/minute	3.241×10^{-4}	kg-calories/min
foot-pounds/minute	2.260×10^{-5}	kw
foot-pounds/second	4.6275	BTU (mean)/hour
foot-pounds/second	0.07717	BTU/minute
foot-pounds/second	0.0018182	horsepower
foot-pounds/second	0.01945	kg-calories/min
foot-pounds/second	0.001356	kilowatts
foot-pounds/second	1.35582	watts (abs)
furlongs	660.0	feet
furlongs	201.17	meters
furlongs	0.125	miles (U.S.)
furlongs	40.0	rods
gallons (Br.)	3.8125×10^{-2}	barrels (U.S.)

Multiply	*by*	*to obtain*
gallons (Br.)	4516.086	cubic centimeters
gallons (Br.)	0.16053	cu ft
gallons (Br.)	277.4	cu inches
gallons (Br.)	1230	drams (U.S. fluid)
gallons (Br.)	4.54596	liters
gallons (Br.)	7.9620×10^4	minims (Br.)
gallons (Br.)	7.3783×10^4	minims (U.S.)
gallons (Br.)	4545.96	mL
gallons (Br.)	1.20094	gallons (U.S.)
gallons (Br.)	160	ounces (Br., fl.)
gallons (Br.)	153.72	ounces (U.S., fl.)
gallons (Br.)	10	pounds (avoirdupois) of water at 62°F
gallons (U.S.)	3.068×10^{-4}	acre-ft
gallons (U.S.)	0.031746	barrels (U.S.)
gallons (U.S.)	3785.434	cubic centimeters
gallons (U.S.)	0.13368	cubic feet (U.S.)
gallons (U.S.)	231	cubic inches
gallons (U.S.)	3.785×10^{-3}	cubic meters
gallons (U.S.)	4.951×10^{-3}	cubic yards
gallons (U.S.)	1024	drams (U.S., fluid)
gallons (U.S.)	0.83268	gallons (Br.)
gallons (U.S.)	0.83267	imperial gal
gallons (U.S.)	3.78533	liters
gallons (U.S.)	6.3950×10^4	minims (Br.)
gallons (U.S.)	6.1440×10^4	minims (U.S.)
gallons (U.S.)	3785	mL
gallons (U.S.)	133.23	ounces (Br., fluid)
gallons (U.S.)	128	ounces (U.S., fluid)
gallons	8	pints (liq.)
gallons	4	quarts (liq.)
gal water (U.S.)	8.345	lb of water
gallons/acre	0.00935	cu m/ha
gallons/day	4.381×10^{-5}	liters/sec
gpd/acre	0.00935	cu m/day/ha
gpd/acre	9.353	liter/day/ha
gallons/capita/day	3.785	liters/capita/day
gpd/cu yd	5.0	L/day/cu m
gpd/ft	0.01242	cu m/day/m
gpd/sq ft	0.0408	cu m/day/sq m
gpd/sq ft	1.698×10^{-5}	cubic meters/hour/sq meter
gpd/sq ft	0.283	cu meter/minute/ha
gpm (gal/min)	8.0208	cfh (cu ft/hr)
gpm	2.228×10^{-3}	cfs (cu ft/sec)

Multiply	*by*	*to obtain*
gpm	4.4021	cubic meters/hr
gpm	0.00144	MGD
gpm	0.0631	liters/sec
gpm/sq ft	2.445	cu meters/hour/sq meter
gpm/sq ft	40.7	L/min/sq meter
gpm/sq ft	0.679	liter/sec/sq meter
gallons/sq ft	40.743	liters/sq meter
gausses (abs)	3.3358×10^{-4}	electrostatic cgs units of magnetic flux density
gausses (abs)	0.99966	gausses (int)
gausses (abs)	1	lines/square centimeter
gausses (abs)	6.452	lines/sq in
gausses (abs)	1	maxwells (abs)/square centimeters
gausses (abs)	6.4516	maxwells (abs)/square inch
gausses (abs)	10^{-8}	webers/sq cm
gausses (abs)	6.452×10^{-8}	webers/sq in
gausses (abs)	10^{-4}	webers/sq meter
gilberts (abs)	0.07958	abampere turns
gilberts (abs)	0.7958	ampere turns
gilberts (abs)	2.998×10^{10}	electrostatic cgs units of magneto motive force
gilberts/cm	0.7958	amp-turns/cm
gilberts/cm	2.021	amp-turns/in
gilberts/cm	79.58	amp-turns/meter
gills (Br.)	142.07	cubic cm
gills (Br.)	5	ounces (British, fluid)
gills (U.S.)	32	drams (fluid)
gills	0.1183	liters
gills	0.25	pints (liq.)
grade	0.01571	radian
grains	0.036571	drams (avoirdupois)
grains	0.01667	drams (troy)
grains (troy)	1.216	grains (avdp)
grains (troy)	0.06480	grams
grains (troy)	6.480×10^{-5}	kilograms
grains (troy)	64.799	milligrams
grains (troy)	2.286×10^{-3}	ounces (avdp)
grains (troy)	2.0833×10^{-3}	ounces (troy)
grains (troy)	0.04167	pennyweights (troy)
grains	1/7000	pounds (avoirdupois)
grains	1.736×10^{-4}	pounds (troy)
grains	6.377×10^{-8}	tons (long)
grains	7.142×10^{-8}	tons (short)
grains/imp gal	14.254	mg/L

Multiply	*by*	*to obtain*
grains/imp. gal	14.254	parts/million (ppm)
grains/U.S. gal	17.118	mg/L
grains/U.S. gal	17.118	parts/million (ppm)
grains/U.S. gal	142.86	lb/mil gal
grams	0.5611	drams (avdp)
grams	0.25721	drams (troy)
grams	980.7	dynes
grams	15.43	grains
grams	9.807×10^{-5}	joules/cm
grams	9.807×10^{-3}	joules/meter (newtons)
grams	10^{-3}	kilograms
grams	10^{3}	milligrams
grams	0.0353	ounces (avdp)
grams	0.03215	ounces (troy)
grams	0.07093	poundals
grams	2.205×10^{-3}	pounds
grams	2.679×10^{-3}	pounds (troy)
grams	9.842×10^{-7}	tons (long)
grams	1.102×10^{-6}	tons (short)
grams-calories	4.1868×10^{7}	ergs
gram-calories	3.0880	foot-pounds
gram-calories	1.5597×10^{-6}	horsepower-hr
gram-calories	1.1630×10^{-6}	kilowatt-hr
gram-calories	1.1630×10^{-3}	watt-hr
gram-calories	3.968×10^{-3}	British Thermal Units (BTU)
gram-calories/sec	14.286	BTU/hr
gram-centimeters	9.2967×10^{-8}	BTU (mean)
gram-centimeters	2.3427×10^{-5}	calories, gram (mean)
gram-centimeters	980.7	ergs
gram-centimeters	7.2330×10^{-5}	foot-pounds
gram-centimeters	9.8067×10^{-5}	joules (abs)
gram-centimeters	2.344×10^{-8}	kilogram-calories
gram-centimeters	10^{-5}	kilogram-meters
gram-centimeters	2.7241×10^{-8}	watt-hours
grams-centimeters^2 (moment of inertia)	2.37305×10^{-6}	pounds-feet^2
grams-centimeters^2 (moment of inertia)	3.4172×10^{-4}	pounds-inch^2
gram-centimeters/second	1.3151×10^{-7}	hp
gram-centimeters/second	9.8067×10^{-8}	kilowatts
gram-centimeters/second	0.065552	lumens
gram-centimeters/second	9.80665×10^{-5}	watt (abs)
grams/cm	5.600×10^{-3}	pounds/inch
grams/cu cm	62.428	pounds/cubic foot
grams/cu cm	0.03613	pounds/cubic inch

Multiply	*by*	*to obtain*
grams/cu cm	8.3454	pounds/gallon (U.S.)
grams/cu cm	3.405×10^{-7}	pounds/mil-foot
grams/cu ft	35.314	grams/cu meter
grams/cu ft	10^6	micrograms/cu ft
grams/cu ft	35.314×10^6	micrograms/cu meter
grams/cu ft	35.3145×10^3	milligrams/cu meter
grams/cu ft	2.2046	pounds/1000 cu ft
grams/cu m	0.43700	grains/cubic foot
grams/cu m	0.02832	grams/cu ft
grams/cu m	28.317×10^3	micrograms/cu ft
grams/cu m	0.06243	pounds/cu ft
grams/liter	58.417	grains/gallon (U.S.)
grams/liter	9.99973×10^{-4}	grams/cubic centimeter
grams/liter	1000	mg/L
grams/liter	1000	parts per million (ppm)
grams/liter	0.06243	pounds/cubic foot
grams/liter	8.345	lb/1000 gal
grams/sq centimeter	2.0481	pounds/sq ft
grams/sq centimeter	0.0142234	pounds/square inch
grams/sq ft	10.764	grams/sq meter
grams/sq ft	10.764×10^3	kilograms/sq km
grams/sq ft	1.0764	milligrams/sq cm
grams/sq ft	10.764×10^3	milligrams/sq meter
grams/sq ft	96.154	pounds/acre
grams/sq ft	2.204	pounds/1000 sq ft
grams/sq ft	30.73	tons/sq mile
grams/sq meter	0.0929	grams/sq ft
grams/sq meter	1000	kilograms/sq km
grams/sq meter	0.1	milligrams/square cm
grams/sq meter	1000	milligrams/sq meter
grams/sq meter	8.921	pounds/acre
grams/sq meter	0.2048	pounds/1000 sq ft
grams/sq meter	2.855	tons/sq mile
g (gravity)	9.80665	meters/sec^2
g (gravity)	32.174	ft/sec^2
hand	10.16	cm
hands	4	inches
hectare (ha)	2.471	acre
hectares	1.076×10^5	sq feet
hectograms	100	grams
hectoliters	100	liters
hectometers	100	meters
hectowatts	100	watts
hemispheres	0.5	spheres

Multiply	*by*	*to obtain*
hemispheres	4	spherical right angles
hemispheres	6.2832	steradians
henries (abs)	10^9	abhenries
henries	1000.0	millihenries
henries (abs)	1.1126×10^{-12}	stathenries
hogsheads (British)	63	gallons (British)
hogsheads (British)	10.114	cubic feet
hogsheads (U.S.)	8.422	cubic feet
hogsheads (U.S.)	0.2385	cubic meters
hogsheads (U.S.)	63	gallons (U.S.)
horsepower	2545.08	BTU (mean)/hour
horsepower	42.44	BTU/min
horsepower	7.457×10^9	erg/sec
horsepower	33,000	ft lb/min
horsepower	550	foot-pounds/second
horsepower	7.6042×10^6	g cm/sec
horsepower, electrical	1.0004	horsepower
horsepower	10.70	kg.-calories/min
horsepower	0.74570	kilowatts (g = 980.665)
horsepower	498129	lumens
horsepower, continental	736	watts (abs)
horsepower, electrical	746	watts (abs)
horsepower (boiler)	9.803	kw
horsepower (boiler)	33.479	BTU/hr
horsepower-hours	2545	BTU (mean)
horsepower-hours	2.6845×10^{13}	ergs
horsepower-hours	6.3705×10^7	ft poundals
horsepower-hours	1.98×10^6	foot-pounds
horsepower-hours	641,190	gram-calories
horsepower-hours	2.684×10^6	joules
horsepower-hours	641.7	kilogram-calories
horsepower-hours	2.737×10^5	kilogram-meters
horsepower-hours	0.7457	kilowatt-hours (abs)
horsepower-hours	26,494	liter atmospheres (normal)
horsepower-hours	745.7	watt-hours
hours	4.167×10^{-2}	days
hours	60	minutes
hours	3600	seconds
hours	5.952×10^{-3}	weeks
hundredweights (long)	112	pounds
hundredweights (long)	0.05	tons (long)
hundredweights (short)	1600	ounces (avoirdupois)
hundredweights (short)	100	pounds
hundredweights (short)	0.0453592	tons (metric)

Multiply	*by*	*to obtain*
hundredweights (short)	0.0446429	tons (long)
inches (British)	2.540	centimeters
inches (U.S.)	2.54000508	centimeters
inches (British)	0.9999972	inches (U.S.)
inches	2.540×10^{-2}	meters
inches	1.578×10^{-5}	miles
inches	25.40	millimeters
inches	10^3	mils
inches	2.778×10^{-2}	yards
inches^2	6.4516×10^{-4}	meter^2
inches^3	1.6387×10^{-5}	meter^3
in. of mercury	0.0334	atm
in. of mercury	1.133	ft of water
in. of mercury (0°C)	13.609	inches of water (60°F)
in. of mercury	0.0345	kgs/square cm
in. of mercury at 32°F	345.31	kilograms/square meter
in. of mercury	33.35	millibars
in. of mercury	25.40	millimeters of mercury
in. of mercury (60°F)	3376.85	newton/meter^2
in. of mercury	70.73	pounds/square ft
in. of mercury at 32°F	0.4912	pounds/square inch
in. of water	0.002458	atmospheres
in. of water	0.0736	in. of mercury
in. of water (at 4°C)	2.540×10^{-3}	kgs/sq cm
in. of water	25.40	kgs/square meter
in. of water (60°F)	1.8663	millimeters of mercury (0°C)
in. of water (60°F)	248.84	newton/meter^2
in. of water	0.5781	ounces/square in
in. of water	5.204	pounds/square ft
in. of water	0.0361	psi
inches/hour	2.54	cm/hr
international ampere	.9998	ampere (absolute)
international volt	1.0003	volts (absolute)
international volt	1.593×10^{-19}	joules (absolute)
international volt	9.654×10^4	joules
joules	9.480×10^{-4}	BTU
joules (abs)	10^7	ergs
joules	23.730	foot poundals
joules (abs)	0.73756	foot-pounds
joules	3.7251×10^{-7}	horsepower hours
joules	2.389×10^{-4}	kg-calories
joules (abs)	0.101972	kilogram-meters
joules	9.8689×10^{-3}	liter atmospheres (normal)
joules	2.778×10^{-4}	watt-hrs

Multiply	*by*	*to obtain*
joules-sec	1.5258×10^{33}	quanta
joules/cm	1.020×10^{4}	grams
joules/cm	10^{7}	dynes
joules/cm	100.0	joules/meter (newtons)
joules/cm	723.3	poundals
joules/cm	22.48	pounds
joules/liter	0.02681	BTU/cu ft
joules/m^2-sec	0.3167	BTU/ft^2-hr
joules/sec	3.41304	BTU/hr
joules/sec	0.056884	BTU/min
joules/sec	1×10^{7}	erg/sec
joules/sec	44.254	ft lb/min
joules/sec	0.73756	ft lb/sec
joules/sec	1.0197×10^{4}	g cm/sec
joules/sec	1.341×10^{-3}	hp
joules/sec	0.01433	kg cal/min
joules/sec	0.001	kilowatts
joules/sec	668	lumens
joules/sec	1	watts
kilograms	564.38	drams (avdp)
kilograms	257.21	drams (troy)
kilograms	980,665	dynes
kilograms	15,432	grains
kilograms	1000	grams
kilograms	0.09807	joules/cm
kilograms	9.807	joules/meter (newtons)
kilograms	1×10^{6}	milligrams
kilograms	35.274	ounces (avdp)
kilograms	32.151	ounces (troy)
kilograms	70.93	poundals
kilograms	2.20462	pounds (avdp)
kilograms	2.6792	pounds (troy)
kilograms	9.84207×10^{-4}	tons (long)
kilograms	0.001	tons (metric)
kilograms	0.0011023	tons (short)
kilogram-calories	3.968	British Thermal Units (BTU)
kilogram-calories	3086	foot-pounds
kilogram-calories	1.558×10^{-3}	horsepower-hours
kilogram-calories	4186	joules
kilogram-calories	426.6	kilogram-meters
kilogram-calories	4.186	kilojoules
kilogram-calories	1.162×10^{-3}	kilowatt-hours
kg-cal/min	238.11	BTU/hr
kg-cal/min	3.9685	BTU/min

Multiply	*by*	*to obtain*
kg-cal/min	6.9770×10^{8}	erg/sec
kg-cal/min	3087.4	ft-lb/min
kg-cal/min	51.457	ft-lb/sec
kg-cal/min	7.1146×10^{5}	g cm/sec
kg-cal/min	0.0936	hp
kg-cal/min	69.769	joules/sec
kg-cal/min	0.0698	kw
kg-cal/min	46636	lumens
kg-cal/min	69.767	watts
kgs-cms. squared	2.373×10^{-3}	pounds-feet squared
kgs-cms. squared	0.3417	pounds-inches squared
kilogram-force (kgf)	9.80665	newton
kilogram-meters	0.0092967	BTU (mean)
kilogram-meters	2.3427	calories, gram (mean)
kilogram-meters	9.80665×10^{7}	ergs
kilogram-meters	232.71	ft poundals
kilogram-meters	7.2330	foot-pounds
kilogram-meters	3.6529×10^{-6}	horsepower-hours
kilogram-meters	9.80665	joules (abs)
kilogram-meters	2.344×10^{-3}	kilogram-calories
kilogram-meters	2.52407×10^{-6}	kilowatt-hours (abs)
kilogram-meters	2.7241×10^{-6}	kilowatt-hours
kilogram-meters	0.096781	liter atmospheres (normal)
kilogram-meters	6.392×10^{-7}	pounds carbon to CO_2
kilogram-meters	9.579×10^{-6}	pounds water evap. at 212°F
kilograms/cubic meter	10^{-3}	grams/cubic cm
kilograms/cubic meter	0.06243	pounds/cubic foot
kilograms/cubic meter	3.613×10^{-5}	pounds/cubic inch
kilograms/cubic meter	3.405×10^{-10}	pounds/mil. foot
kilograms/m^3-day	0.0624	lb/cu ft-day
kilograms/cu meter-day	62.43	pounds/1000 cu ft-day
kilograms/ha	0.8921	pounds/acre
kilograms/meter	0.6720	pounds/foot
kilograms/sq cm	980,665	dynes
kilograms/sq cm	0.96784	atmosphere
kilograms/sq cm	32.81	feet of water
kilograms/sq cm	28.96	inches of mercury
kilograms/sq cm	735.56	mm of mercury
kilograms/sq cm	2048	pounds/sq ft
kilograms/sq cm	14.22	pounds/square inch
kilograms/sq km	92.9×10^{-6}	grams/sq ft
kilograms/sq km	0.001	grams/sq meter
kilograms/sq km	0.0001	milligrams/sq cm
kilograms/sq km	1.0	milligrams/sq meter

Multiply	*by*	*to obtain*
kilograms/sq km	8.921×10^{-3}	pounds/acre
kilograms/sq km	204.8×10^{-6}	pounds/1000 sq ft
kilograms/sq km	2.855×10^{-3}	tons/sq mile
kilograms/sq meter	9.6784×10^{-5}	atmospheres
kilograms/sq meter	98.07×10^{-6}	bars
kilograms/sq meter	98.0665	dynes/sq centimeters
kilograms/sq meter	3.281×10^{-3}	feet of water at 39.2°F
kilograms/sq meter	0.1	grams/sq centimeters
kilograms/sq meter	2.896×10^{-3}	inches of mercury at 32°F
kilograms/sq meter	0.07356	mm of mercury at 0°C
kilograms/sq meter	0.2048	pounds/square foot
kilograms/sq meter	0.00142234	pounds/square inch
kilograms/sq mm.	10^6	kg/square meter
kilojoule	0.947	BTU
kilojoules/kilogram	0.4295	BTU/pound
kilolines	1000.0	maxwells
kiloliters	10^3	liters
kilometers	10^5	centimeters
kilometers	3281	feet
kilometers	3.937×10^4	inches
kilometers	10^3	meters
kilometers	0.53961	miles (nautical)
kilometers	0.6214	miles (statute)
kilometers	10^6	millimeters
kilometers	1093.6	yards
kilometers/hr	27.78	cm/sec
kilometers/hr	54.68	feet/minute
kilometers/hr	0.9113	ft/sec
kilometers/hr	0.5396	knot
kilometers/hr	16.67	meters/minute
kilometers/hr	0.2778	meters/sec
kilometers/hr	0.6214	miles/hour
kilometers/hour/sec	27.78	cms/sec/sec
kilometers/hour/sec	0.9113	ft/sec/sec
kilometers/hour/sec	0.2778	meters/sec/sec
kilometers/hour/sec	0.6214	miles/hr/sec
kilometers/min	60	kilometers/hour
kilonewtons/sq m	0.145	psi
kilowatts	56.88	BTU/min
kilowatts	4.425×10^4	foot-pounds/min
kilowatts	737.6	ft-lb/sec
kilowatts	1.341	horsepower
kilowatts	14.34	kg-cal/min
kilowatts	10^3	watts

Multiply	*by*	*to obtain*
kilowatt-hrs	3413	BTU (mean)
kilowatt-hrs	3.600×10^{13}	ergs
kilowatt-hrs	2.6552×10^{6}	foot-pounds
kilowatt-hrs	859,850	gram-calories
kilowatt-hrs	1.341	horsepower hours
kilowatt-hrs	3.6×10^{6}	joules
kilowatt-hrs	860.5	kg-calories
kilowatt-hrs	3.6709×10^{5}	kilogram-meters
kilowatt-hrs	3.53	pounds of water evaporated from and at 212°F
kilowatt-hrs	22.75	pounds of water raised from 62° to 212°F
knots	6080	feet/hr
knots	1.689	feet/sec
knots	1.8532	kilometers/hr
knots	0.5144	meters/sec
knots	1.0	miles (nautical)/hour
knots	1.151	miles (statute)/hour
knots	2,027	yards/hr
lambert	2.054	candle/in^2
lambert	929	footlambert
lambert	0.3183	stilb
langley	1	15° gram-calorie/cm^2
langley	3.6855	BTU/ft^2
langley	0.011624	Int. kw-hr/m^2
langley	4.1855	joules (abs)/cm^2
leagues (nautical)	3	miles (nautical)
leagues (statute)	3	miles (statute)
light years	63,274	astronomical units
light years	9.4599×10^{12}	kilometers
light years	5.8781×10^{12}	miles
lignes (Paris lines)	1/12	ponces (Paris inches)
lines/sq cm	1.0	gausses
lines/sq in	0.1550	gausses
lines/sq in	1.550×10^{-9}	webers/sq cm
lines/sq in	10^{-8}	webers/sq in
lines/sq in	1.550×10^{-5}	webers/sq meter
links (engineer's)	12.0	inches
links (Gunter's)	0.01	chains (Gunter's)
links (Gunter's)	0.66	feet
links (Ramden's)	0.01	chains (Ramden's)
links (Ramden's)	1	feet
links (surveyor's)	7.92	inches
liters	8.387×10^{-3}	barrels (U.S.)

Multiply	*by*	*to obtain*
liters	0.02838	bushels (U.S. dry)
liters	1000.028	cubic centimeters
liters	0.035316	cubic feet
liters	61.025	cu inches
liters	10^{-3}	cubic meters
liters	1.308×10^{-3}	cubic yards
liters	270.5179	drams (U.S. fl)
liters	0.21998	gallons (Br.)
liters	0.26417762	gallons (U.S.)
liters	16,894	minims (Br.)
liters	16,231	minims (U.S.)
liters	35.196	ounces (Br. fl)
liters	33.8147	ounces (U.S. fl)
liters	2.113	pints (liq.)
liters	1.0566828	quarts (U.S. liq.)
liter-atmospheres (normal)	0.096064	BTU (mean)
liter-atmospheres (normal)	24.206	calories, gram (mean)
liter-atmospheres (normal)	1.0133×10^{9}	ergs
liter-atmospheres (normal)	74.735	foot-pounds
liter-atmospheres (normal)	3.7745×10^{-5}	horsepower hours
liter-atmospheres (normal)	101.33	joules (abs)
liter-atmospheres (normal)	10.33	kilogram-meters
liter-atmospheres (normal)	2.4206×10^{-2}	kilogram calories
liter-atmospheres (normal)	2.815×10^{-5}	kilowatt-hours
liter/cu m-sec	60.0	cfm/1000 cu ft
liters/minute	5.885×10^{-4}	cubic feet/sec
liters/minute	4.403×10^{-3}	gallons/sec
liter/person-day	0.264	gpcd
liters/sec	2.119	cu ft /min
liters/sec	3.5316×10^{-2}	cu ft /sec
liters/sec	15.85	gallons/minute
liters/sec	0.02282	MGD
$\log_{10} N$	2.303	$\log_e N$ or ln N
$\log_e N$ or ln N	0.4343	$\log_{10} N$
lumens	0.07958	candle-power (spherical)
lumens	0.00147	watts of maximum visibility radiation
lumens/sq. centimeters	1	lamberts
lumens/sq cm/steradian	3.1416	lamberts
lumens/sq ft	1	foot-candles
lumens/sq ft	10.764	lumens/sq meter
lumens/sq ft/steradian	3.3816	millilamberts
lumens/sq meter	0.09290	foot-candles or lumens/sq
lumens/sq meter	10^{-4}	phots
lux	0.09290	foot-candles

Multiply	*by*	*to obtain*
lux	1	lumens/sq meter
lux	10^{-4}	phots
maxwells	0.001	kilolines
maxwells	10^{-8}	webers
megajoule	0.3725	horsepower-hour
megalines	10^{6}	maxwells
megohms	10^{12}	microhms
megohms	10^{6}	ohms
meters	10^{10}	angstrom units
meters	100	centimeters
meters	0.5467	fathoms
meters	3.280833	feet (U.S.)
meters	39.37	inches
meters	10^{-3}	kilometers
meters	5.396×10^{-4}	miles (naut.)
meters	6.2137×10^{-4}	miles (statute)
meters	10^{3}	millimeters
meters	10^{9}	millimicrons
meters	1.09361	yards (U.S.)
meters	1.179	varas
meter-candles	1	lumens/sq meter
meter-kilograms	9.807×10^{7}	centimeter-dynes
meter-kilograms	10^{5}	centimeter-grams
meter-kilograms	7.233	pound-feet
meters/minute	1.667	centimeters/sec
meters/minute	3.281	feet/minute
meters/minute	0.05468	feet/second
meters/minute	0.06	kilograms/hour
meters/minute	0.03238	knots
meters/minute	0.03728	miles/hour
meters/second	196.8	feet/minute
meters/second	3.281	feet/second
meters/second	3.6	kilometers/hour
meters/second	0.06	kilometers/min
meters/second	1.944	knots
meters/second	2.23693	miles/hour
meters/second	0.03728	miles/minute
meters/sec/sec	100.0	cm/sec/sec
meters/sec/sec	3.281	feet/sec/sec
meters/sec/sec	3.6	km/hour/sec
meters/sec/sec	2.237	miles/hour/sec
microfarad	10^{-6}	farads
micrograms	10^{-6}	grams
micrograms/cu ft	10^{-6}	grams/cu ft

Multiply	*by*	*to obtain*
micrograms/cu ft	35.314×10^{-6}	grams/cu m
micrograms/cu ft	35.314	microgram/cu m
micrograms/cu ft	35.314×10^{-3}	milligrams/cu m
micrograms/cu ft	2.2046×10^{-6}	pounds/1000 cu ft
micrograms/cu m	28.317×10^{-9}	grams/cu ft
micrograms/cu m	10^{-6}	grams/ cu m
micrograms/cu m	0.02832	micrograms/cu ft
micrograms/cu m	0.001	milligrams/cu m
micrograms/cu m	62.43×10^{-9}	pounds/1000 cu ft
micrograms/cu m	$\frac{0.02404}{\text{molecular weight of gas}}$	ppm by volume (20°C)
micrograms/cu m	834.7×10^{-6}	ppm by weight
micrograms/liter	1000.0	micrograms/cu m
micrograms/liter	1.0	milligrams/cu m
micrograms/liter	62.43×10^{-9}	pounds/cu ft
micrograms/liter	$\frac{24.04}{\text{molecular weight of gas}}$	ppm by volume (20°C)
micrograms/liter	0.834.7	ppm by weight
microhms	10^{-12}	megohms
microhms	10^{-6}	ohms
microliters	10^{-6}	liters
microns	10^{4}	angstrom units
microns	1×10^{-4}	centimeters
microns	3.9370×10^{-5}	inches
microns	10^{-6}	meters
miles (naut.)	6080.27	feet
miles (naut.)	1.853	kilometers
miles (naut.)	1.853	meters
miles (naut.)	1.1516	miles (statute)
miles (naut.)	2027	yards
miles (statute)	1.609×10^{5}	centimeters
miles (statute)	5280	feet
miles (statute)	6.336×10^{4}	inches
miles (statute)	1.609	kilometers
miles (statute)	1609	meters
miles (statute)	0.8684	miles (naut.)
miles (statute)	320	rods
miles (statute)	1760	yards
miles/hour	44.7041	centimeter/second
miles/hour	88	feet/min
miles/hour	1.4667	feet/sec
miles/hour	1.6093	kilometers/hour
miles/hour	0.02682	km/min

Multiply	*by*	*to obtain*
miles/hour	0.86839	knots
miles/hour	26.82	meters/min
miles/hour	0.447	meters/sec
miles/hour	0.1667	miles/min
miles/hour/sec	44.70	cms/sec/sec
miles/hour/sec	1.4667	ft/sec/sec
miles/hour/sec	1.6093	km/hour/sec
miles/hour/sec	0.4470	m/sec/sec
miles/min	2682	centimeters/sec
miles/min	88	ft/sec
miles/min	1.609	km/min
miles/min	0.8684	knots/min
miles/min	60	miles/hour
miles-feet	9.425×10^{-6}	cu inches
millibars	0.00987	atmospheres
millibars	0.30	inches of mercury
millibars	0.75	millimeters of mercury
milliers	10^3	kilograms
millimicrons	1×10^{-9}	meters
milligrams	0.01543236	grains
milligrams	10^{-3}	grams
milligrams	10^{-6}	kilograms
milligrams	3.5274×10^{-5}	ounces (avdp)
milligrams	2.2046×10^{-6}	pounds (avdp)
milligrams/assay ton	1	ounces (troy)/ton (short)
milligrams/cu m	283.2×10^{-6}	grams/cu ft
milligrams/cu m	0.001	grams/cu m
milligrams/cu m	1000.0	micrograms/cu m
milligrams/cu m	28.32	micrograms/cu ft
milligrams/cu m	1.0	micrograms/liter
milligrams/cu m	62.43×10^{-6}	pounds/1000 cu ft
milligrams/cu m	$\frac{24.04}{\text{molecular weight of gas}}$	ppm by volume (20°C)
milligrams/cu m	0.8347	ppm by weight
milligrams/joule	5.918	pounds/horsepower-hour
milligrams/liter	0.05841	grains/gallon
milligrams/liter	0.07016	grains/imp. gal
milligrams/liter	0.0584	grains/U.S. gal
milligrams/liter	1.0	parts/million
milligrams/liter	8.345	lb/mil gal
milligrams/sq cm	0.929	grams/sq ft
milligrams/sq cm	10.0	grams/sq meter
milligrams/sq cm	10^4	kilograms/sq km

Multiply	*by*	*to obtain*
milligrams/sq cm	10^4	milligrams/sq meter
milligrams/sq cm	2.048	pounds/1000 sq ft
milligrams/sq cm	89.21	pounds/acre
milligrams/sq cm	28.55	tons/sq mile
milligrams/sq meter	92.9×10^{-6}	grams/sq ft
milligrams/sq meter	0.001	grams/sq meter
milligrams/sq meter	1.0	kilograms/sq km
milligrams/sq meter	0.0001	milligrams/sq cm
milligrams/sq meter	8.921×10^{-3}	pounds/acre
milligrams/sq meter	204.8×10^{-6}	pounds/1000 sq ft
milligrams/sq meter	2.855×10^{-3}	tons/sq mile
millihenries	0.001	henries
milliters	1	cubic centimeters
milliliters	3.531×10^{-5}	cu ft
milliliters	6.102×10^{-2}	cu in
milliliters	10^{-6}	cu m
milliliters	2.642×10^{-4}	gal (U.S.)
milliliters	10^{-3}	liters
milliliters	0.03381	ounces (U.S. fl)
millimeters	0.1	centimeters
millimeters	3.281×10^{-3}	feet
millimeters	0.03937	inches
millimeters	10^{-6}	kilometers
millimeters	0.001	meters
millimeters	6.214×10^{-7}	miles
millimeters	39.37	mils
millimeters	1.094×10^{-3}	yards
millimeters of mercury	1.316×10^{-3}	atmospheres
millimeters of mercury	0.0394	inches of mercury
millimeters of mercury (0°C)	0.5358	inches of water (60°F)
millimeters of mercury	1.3595×10^{-3}	kg/sq cm
millimeter of mercury (0°C)	133.3224	newton/meter^2
millimeters of mercury	0.01934	pounds/sq in
millimeters/sec	11.81	feet/hour
million gallons	306.89	acre-ft
million gallons	3785.0	cubic meters
million gallons	3.785	mega liters (1×10^6)
million gallons/day (MGD)	1.547	cu ft/sec
MGD	3785	cu m/day
MGD	0.0438	cubic meters/sec
MGD	43.808	liters/sec
MGD/acre	9360	cu m/day/ha
MGD/acre	0.039	cu meters/hour/sq meter

Multiply	*by*	*to obtain*
mils	0.002540	centimeters
mils	8.333×10^{-5}	feet
mils	0.001	inches
mils	2.540×10^{-8}	kilometers
mils	25.40	microns
mils	2.778×10^{-5}	yards
miner's in.	1.5	cu ft/min
miner's inches (Ariz., Calif. Mont., and Ore.)	0.025	cubic feet/second
miner's in. (Colorado)	0.02604	cubic feet/second
miner's inches (Idaho, Kan., Neb., Nev., N. Mex., N. Dak., S. Dak. and Utah)	0.020	cubic feet/second
minims (British)	0.05919	cubic centimeter
minims (U.S.)	0.06161	cubic centimeters
minutes (angles)	0.01667	degrees
minutes (angles)	1.852×10^{-4}	quadrants
minutes (angles)	2.909×10^{-4}	radians
minutes (angle)	60	seconds (angle)
months (mean calendar)	30.4202	days
months (mean calendar)	730.1	hours
months (mean calendar)	43805	minutes
months (mean calendar)	2.6283×10^{6}	seconds
myriagrams	10	kilograms
myriameters	10	kilometers
myriawatts	10	kilowatts
nepers	8.686	decibels
newtons	10^{5}	dynes
newtons	0.10197	kilograms
newtons	0.22481	pounds
newtons/sq meter	1.00	pascals (Pa)
noggins (British)	1/32	gallons (British)
No./cu.cm.	28.316×10^{3}	No./cu ft
No./cu.cm.	10^{6}	No./cu meter
No./cu.cm.	1000.0	No./liter
No./cu.ft.	35.314×10^{-6}	No./cu cm
No./cu.ft.	35.314	No./cu meter
No./cu.ft.	35.314×10^{-3}	No./liter
No./cu. meter	10^{-6}	No./cu cm
No./cu. meter	28.317×10^{-3}	No./cu ft
No./cu. meter	0.001	No./liter
No./liter	0.001	No./cu cm
No./liter	28.316	No./cu ft

Multiply	*by*	*to obtain*
No./liter	1000.0	No./cu meter
oersteds (abs)	1	electromagnetic cgs units of magnetizing force
oersteds (abs)	2.9978×10^{10}	electrostatic cgs units of magnetizing force
ohms	10^{9}	abohms
ohms	1.1126×10^{-12}	statohms
ohms	10^{-6}	megohms
ohms	10^{6}	microhms
ohms (International)	1.0005	ohms (absolute)
ounces (avdp)	16	drams (avoirdupois)
ounces (avdp)	7.2917	drams (troy)
ounces (avdp)	437.5	grains
ounces (avdp)	28.349527	grams
ounces (avdp)	0.028350	kilograms
ounces (avdp)	2.8350×10^{4}	milligrams
ounces (avdp)	0.9114583	ounces (troy)
ounces (avdp)	0.0625	pounds (avoirdupois)
ounces (avdp)	0.075955	pounds (troy)
ounces (avdp)	2.790×10^{-5}	tons (long)
ounces (avdp)	2.835×10^{-5}	tons (metric)
ounces (avdp)	3.125×10^{-5}	tons (short)
ounces (Br. fl)	2.3828×10^{-4}	barrels (U.S.)
ounces (Br. fl)	1.0033×10^{-3}	cubic feet
ounces (Br. fl)	1.73457	cubic inches
ounces (Br. fl)	7.6860	drams (U.S. fl)
ounces (Br. fl)	6.250×10^{-3}	gallons (Br.)
ounces (Br. fl)	0.07506	gallons (U.S.)
ounces (Br. fl)	2.84121×10^{-2}	liters
ounces (Br. fl)	480	minims (Br.)
ounces (Br. fl)	461.160	minims (U.S.)
ounces (Br. fl)	28.4121	mL
ounces (Br. fl)	0.9607	ounces (U.S. fl)
ounces (troy)	17.554	drams (avdp)
ounces (troy)	8	drams (troy)
ounces (troy)	480	grains (troy)
ounces (troy)	31.103481	grams
ounces (troy)	0.03110	kilograms
ounces (troy)	1.09714	ounces (avoirdupois)
ounces (troy)	20	pennyweights (troy)
ounces (troy)	0.068571	pounds (avdp)
ounces (troy)	0.08333	pounds (troy)
ounces (troy)	3.061×10^{-5}	tons (long)
ounces (troy)	3.429×10^{-5}	tons (short)

Multiply	*by*	*to obtain*
ounces (U.S. fl)	2.48×10^{-4}	barrels (U.S.)
ounces (U.S. fl)	29.5737	cubic centimeters
ounces (U.S. fl)	1.0443×10^{-3}	cubic feet
ounces (U.S. fl)	1.80469	cubic inches
ounces (U.S. fl)	8	drams (fluid)
ounces (U.S. fl)	6.5053×10^{-3}	gallons (Br.)
ounces (U.S. fl)	7.8125×10^{-3}	gallons (U.S.)
ounces (U.S. fl)	29.5729	milliliters
ounces (U.S. fl)	499.61	minims (Br.)
ounces (U.S. fl)	480	minims (U.S.)
ounces (U.S. fl)	1.0409	ounces (Br. fl)
ounces/sq inch	4309	dynes/sq cm
ounces/sq. inch	0.0625	pounds/sq inch
paces	30	inches
palms (British)	3	inches
parsecs	3.260	light years
parsecs	3.084×10^{13}	kilometers
parsecs	3.084×10^{16}	meters
parsec	19×10^{12}	miles
parts/billion (ppb)	10^{-3}	mg/L
parts/million (ppm)	0.07016	grains/imp. gal.
parts/million	0.058417	grains/gallon (U.S.)
parts/million	1.0	mg/liter
parts/million	8.345	lbs/million gallons
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{24.04}$	micrograms/liter
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{0.02404}$	micrograms/cu meter
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{24.04}$	milligrams/cu meter
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{28.8}$	ppm by weight
ppm by volume (20°C)	$\frac{\text{molecular weight of gas}}{385.1 \times 10^{6}}$	pounds/cu ft
ppm by weight	1.198×10^{-3}	micrograms/cu meter
ppm by weight	1.198	micrograms/liter
ppm by weight	1.198	milligrams/cu meter
ppm by weight	$\frac{28.8}{\text{molecular weight of gas}}$	ppm by volume (20°C)
ppm by weight	7.48×10^{-6}	pounds/cu ft
pecks (British)	0.25	bushels (British)
pecks (British)	554.6	cubic inches

Multiply	*by*	*to obtain*
pecks (British)	9.091901	liters
pecks (U.S.)	0.25	bushels (U.S.)
pecks (U.S.)	537.605	cubic inches
pecks (U.S.)	8.809582	liters
pecks (U.S.)	8	quarts (dry)
pennyweights	24	grains
pennyweights	1.555174	grams
pennyweights	0.05	ounces (troy)
pennyweights (troy)	4.1667×10^{-3}	pounds (troy)
perches (masonry)	24.75	cubic feet
phots	929.0	foot-candles
phots	1	lumen incident/sq cm
phots	10^4	lux
picas (printers')	1/6	inches
pieds (French feet)	0.3249	meters
pints (dry)	33.6003	cubic inches
pints (liq.)	473.179	cubic centimeters
pints (liq.)	0.01671	cubic feet
pints (liq.)	4.732×10^{-4}	cubic meters
pints (liq.)	6.189×10^{-4}	cubic yards
pints (liq.)	0.125	gallons
pints (liq.)	0.4732	liters
pints (liq.)	16	ounces (U.S. fluid)
pints (liq.)	0.5	quarts (liq.)
planck's constant	6.6256×10^{-27}	erg-seconds
poise	1.00	gram/cm sec
poise	0.1	newton-second/meter^2
population equivalent (PE)	0.17	pounds BOD
pottles (British)	0.5	gallons (British)
pouces (Paris inches)	0.02707	meters
pouces (Paris inches)	0.08333	pieds (Paris feet)
poundals	13,826	dynes
poundals	14.0981	grams
poundals	1.383×10^{-3}	joules/cm
poundals	0.1383	joules/meter (newton)
poundals	0.01410	kilograms
poundals	0.031081	pounds
pounds (avdp)	256	drams (avdp)
pounds (avdp)	116.67	drams (troy)
pounds (avdp)	444,823	dynes
pounds (avdp)	7000	grains
pounds (avdp)	453.5924	grams
pounds (avdp)	0.04448	joules/cm
pounds (avdp)	4.448	joules/meter (newtons)

Multiply	*by*	*to obtain*
pounds (avdp)	0.454	kilograms
pounds (avdp)	4.5359×10^5	milligrams
pounds (avdp)	16	ounces (avdp)
pounds (avdp)	14.5833	ounces (troy)
pounds (avdp)	32.17	poundals
pounds (avdp)	1.2152778	pounds (troy)
pounds (avdp)	4.464×10^{-4}	tons (long)
pounds (avdp)	0.0005	tons (short)
pounds (troy)	210.65	drams (avdp)
pounds (troy)	96	drams (troy)
pounds (troy)	5760	grains
pounds (troy)	373.2418	grams
pounds (troy)	0.37324	kilograms
pounds (troy)	3.7324×10^5	milligrams
pounds (troy)	13.1657	ounces (avdp)
pounds (troy)	12.0	ounces (troy)
pounds (troy)	240.0	pennyweights (troy)
pounds (troy)	0.8229	pounds (avdp)
pounds (troy)	3.6735×10^{-4}	tons (long)
pounds (troy)	3.7324×10^{-4}	tons (metric)
pounds (troy)	4.1143×10^{-4}	tons (short)
pounds (avdp)-force	4.448	newtons
pounds-force-sec/ft^2	47.88026	newton-sec/meter^2
pounds (avdp)-mass	0.4536	kilograms
pounds-mass/ft^3	16.0185	kilogram/meter^3
pounds-mass/ft-sec	1.4882	mewton-sec/meter^2
pounds of BOD	5.882	population equivalent (PE)
pounds of carbon to CO_2	14,544	BTU (mean)
pounds of water	0.0160	cu ft
pounds of water	27.68	cu in
pounds of water	0.1198	gallons
pounds of water evaporated at 212°F	970.3	BTU
pounds of water per min	2.699×10^{-4}	cubic feet/sec
pound-feet	13,825	centimeter-grams
pound-feet (torque)	1.3558×10^7	dyne-centimeters
pound-feet	0.1383	meter-kilograms
pounds-feet squared	421.3	kg-cm squared
pounds-feet squared	144	pounds-inches squared
pounds-inches squared	2926	kg-cm squared
pounds-inches squared	6.945×10^{-3}	pounds-feet squared
pounds/acre	0.0104	grams/sq ft
pounds/acre	0.1121	grams/sq meter
pounds/acre	1.121	kg/ha
pounds/acre	112.1	kilograms/sq km

Multiply	*by*	*to obtain*
pounds/acre	0.01121	milligrams/sq cm
pounds/acre	112.1	milligrams/sq meter
pounds/acre	0.023	pounds/1000 sq ft
pounds/acre	0.32	tons/sq mile
pounds/acre/day	0.112	g/day/sq m
pounds/cu ft	0.0160	g/mL
pounds/cu ft	16.02	kg/cu m
pounds/cu ft	16.018×10^9	micrograms/cu meter
pounds/cu ft	16.018×10^6	micrograms/liter
pounds/cu ft	16.018×10^6	milligrams/cu meter
pounds/cu ft	$\frac{385.1 \times 10^6}{\text{molecular weight of gas}}$	ppm by volume (20°C)
pounds/cu ft	133.7×10^3	ppm by weight
pounds/cu ft	5.787×10^{-4}	lb/cu in
pounds/cu ft	5.456×10^{-9}	pounds/mil-foot
pounds/1000 cu ft	0.35314	grams/cu ft
pounds/1000 cu ft	16.018	grams/cu m
pounds/1000 cu ft	353.14×10^3	micrograms/cu ft
pounds/1000 cu ft	16.018×10^6	microgram/cu m
pounds/1000 cu ft	16.018×10^3	milligrams/cu m
pounds/cubic inch	27.68	grams/cubic cm
pounds/cubic inch	2.768×10^4	kgs/cubic meter
pounds/cubic inch	1728	pounds/cubic foot
pounds/cubic inch	9.425×10^{-6}	pounds/mil foot
pounds/day/acre-ft	3.68	g/day/cu m
pounds/day/cu ft	16	kg/day/cu m
pounds/day/cu yd	0.6	kg/day/cu m
pounds/day/sq ft	4,880	g/day/sq m
pounds/ft	1.488	kg/m
pounds/gal	$454\,\text{g}/3.7851\text{L} = 119.947$	g/liter
pounds/1000-gal	120	g/1000-liters
pounds/horsepower-hour	0.169	mg/joule
pounds/in	178.6	g/cm
pounds/mil-foot	2.306×10^6	gms/cu cm
pounds/mil gal	0.12	g/cu m
pounds/sq ft	4.725×10^{-4}	atmospheres
pounds/sq ft	0.01602	ft of water
pounds/sq ft	0.01414	inches of mercury
pounds/sq ft	4.8824×10^{-4}	kgs/sq cm
pounds/sq ft	4.88241	kilograms/square meter
pounds/sq ft	47.9	newtons/sq m
pounds/sq ft	6.944×10^{-3}	pounds/sq inch
pounds/1000 sq ft	0.4536	grams/sq ft

Multiply	*by*	*to obtain*
pounds/1000 sq ft	4.882	grams/sq meter
pounds/1000 sq ft	4882.4	kilograms/sq km
pounds/1000 sq ft	0.4882	milligrams/sq cm
pounds/1000 sq ft	4882.4	milligrams/sq meter
pounds/1000 sq ft	43.56	pounds/acre
pounds/1000 sq ft	13.94	tons/sq mile
pounds/sq in	0.068046	atmospheres
pounds/sq in	2.307	ft of water
pounds/sq in	70.307	grams/square centimeter
pounds/sq in	2.036	in of mercury
pounds/sq in	0.0703	kgs/square cm
pounds/sq in	703.07	kilograms/square meter
pounds/sq in	51.715	millimeters of mercury
pounds/sq in	6894.76	newton/meter^2
pounds/sq in	51.715	millimeters of mercury at 0°C
pounds/sq in	144	pounds/sq foot
pounds/sq in (abs)	1	pound/sq in (gage) + 14.696
proof (U.S.)	0.5	percent alcohol by volume
puncheons (British)	70	gallons (British)
quadrants (angle)	90	degrees
quadrants (angle)	5400	minutes
quadrants (angle)	3.24×10^5	seconds
quadrants (angle)	1.571	radians
quarts (dry)	67.20	cubic inches
quarts (liq.)	946.4	cubic centimeters
quarts (liq.)	0.033420	cubic feet
quarts (liq.)	57.75	cubic inches
quarts (liq.)	9.464×10^{-4}	cubic meters
quarts (liq.)	1.238×10^{-3}	cubic yards
quarts (liq.)	0.25	gallons
quarts (liq.)	0.9463	liters
quarts (liq.)	32	ounces (U.S., fl)
quarts (liq.)	0.832674	quarts (British)
quintals (long)	112	pounds
quintals (metric)	100	kilograms
quintals (short)	100	pounds
quires	24	sheets
radians	57.29578	degrees
radians	3438	minutes
radians	0.637	quadrants
radians	2.063×10^5	seconds
radians/second	57.30	degrees/second
radians/second	9.549	revolutions/min
radians/second	0.1592	revolutions/sec

Multiply	*by*	*to obtain*
radians/sec/sec	573.0	revs/min/min
radians/sec/sec	9.549	revs/min/sec
radians/sec/sec	0.1592	revs/sec/sec
reams	500	sheets
register tons (British)	100	cubic feet
revolutions	360	degrees
revolutions	4	quadrants
revolutions	6.283	radians
revolutions/minute	6	degrees/second
revolutions/minute	0.10472	radians/second
revolutions/minute	0.01667	revolutions/sec
revolutions/minute^2	0.0017453	radians/sec/sec
revs/min/min	0.01667	revs/min/sec
revs/min/min	2.778×10^{-4}	revs/sec/sec
revolutions/second	360	degrees/second
revolutions/second	6.283	radians/second
revolutions/second	60	revs/minute
revs/sec/sec	6.283	rads/sec/sec
revs/sec/sec	3600	revs/min/min
revs/sec/sec	60	revs/min/sec
reyns	6.8948×10^{6}	centipoises
rod	.25	chain (gunters)
rods	16.5	feet
rods	5.0292	meters
rods	3.125×10^{-3}	miles
rods (surveyors' means)	5.5	yards
roods (British)	0.25	acres
scruples	1/3	drams (troy)
scruples	20	grains
sections	1	square miles
seconds (mean solar)	1.1574×10^{-5}	days
seconds (angle)	2.778×10^{-4}	degrees
seconds (mean solar)	2.7778×10^{-4}	hours
seconds (angle)	0.01667	minutes
seconds (angle)	3.087×10^{-6}	quadrants
seconds (angle)	4.848×10^{-6}	radians
slugs	14.59	kilogram
slugs	32.174	pounds
space, entire (solid angle)	12.566	steradians
spans	9	inches
spheres (solid angle)	12.57	steradians
spherical right angles	0.25	hemispheres
spherical right angles	0.125	spheres
spherical right angles	1.571	steradians

Multiply	*by*	*to obtain*
square centimeters	1.973×10^{5}	circular mils
square centimeters	1.07639×10^{-3}	square feet (U.S.)
square centimeters	0.15499969	square inches (U.S.)
square centimeters	10^{-4}	square meters
square centimeters	3.861×10^{-11}	square miles
square centimeters	100	square millimeters
square centimeters	1.196×10^{-4}	square yards
square centimeters-square centimeter (moment of area)	0.024025	square inch-square inch
square chains (gunter's)	0.1	acres
square chains (gunter's)	404.7	square meters
square chains (Ramden's)	0.22956	acres
square chains (Ramden's)	10000	square feet
square feet	2.29×10^{-5}	acres
square feet	1.833×10^{8}	circular mils
square feet	144	square inches
square feet	0.092903	square meters
square feet	929.0341	square centimeters
square feet	3.587×10^{-8}	square miles
square feet	1/9	square yards
square feet/cu ft	3.29	sq m/cu m
square foot-square foot (moment of area)	20,736	square inch-square inch
square inches	1.273×10^{6}	circular mils
square inches	6.4516258	square centimeters
square inches	6.944×10^{-3}	square feet
square inches	645.2	square millimeters
square inches	10^{6}	square mils
square inches	7.71605×10^{-4}	square yards
square inches-inches sqd.	41.62	sq cm-cm sqd
square inches-inches sqd.	4.823×10^{-5}	sq feet-feet sqd
square kilometers	247.1	acres
square kilometers	10^{10}	square centimeters
square kilometers	10.76×10^{6}	square feet
square kilometers	1.550×10^{9}	square inches
square kilometers	10^{6}	square meters
square kilometers	0.3861006	square miles (U.S.)
square kilometers	1.196×10^{6}	square yards
square links (Gunter's)	10^{-5}	acres (U.S.)
square links (Gunter's)	0.04047	square meters
square meters	2.471×10^{-4}	acres (U.S.)
square meters	10^{4}	square centimeters
square meters	10.76387	square feet (U.S.)
square meters	1550	square inches

Multiply	*by*	*to obtain*
square meters	3.8610×10^{-7}	square miles (statute)
square meters	10^{6}	square millimeters
square meters	1.196	square yards (U.S.)
square miles	640	acres
square miles	2.78784×10^{7}	square feet
square miles	2.590	sq km
square miles	2.5900×10^{6}	square meters
square miles	3.098×10^{6}	square yards
square millimeters	1.973×10^{3}	circular mils
square millimeters	0.01	square centimeters
square millimeters	1.076×10^{-5}	square feet
square millimeters	1.550×10^{-3}	square inches
square mils	1.273	circular mils
square mils	6.452×10^{-6}	square centimeters
square mils	10^{-6}	square inches
square rods	272.3	square feet
square yard	2.1×10^{-4}	acres
square yards	8361	square centimeters
square yards	9	square feet
square yards	1296	square inches
square yards	0.8361	square meters
square yards	3.228×10^{-7}	square miles
square yards	8.361×10^{5}	square millimeters
statamperes	3.33560×10^{-10}	amperes (abs)
statcoulombs	3.33560×10^{-10}	coulombs (abs)
statcoulombs/kilogram	1.0197×10^{-6}	statcoulombs/dyne
statfarads	1.11263×10^{-12}	farads (abs)
stathenries	8.98776×10^{11}	henries (abs)
statohms	8.98776×10^{11}	ohms (abs)
statvolts	299.796	volts (abs)
statvolts/inch	118.05	volts (abs)/centimeter
statwebers	2.99796×10^{10}	electromagnetic cgs units of magnetic flux
statwebers	1	electrostatic cgs units of magnetic flux
stilb	2919	footlambert
stilb	1	int. candle cm^{-2}
stilb	3.142	lambert
stoke (kinematic viscosity)	10^{-4}	$meter^2$/second
stones (British)	6.350	kilograms
stones (British)	14	pounds
temp. (degs. C.) + 273	1	abs. temp. (degs. K.)
temps (degs. C.) + 17.8	1.8	temp. (degs. Fahr.)
temps. (degs. F.) + 460	1	abs. temp. (degs. R.)
temps. (degs. F.) − 32	5/9	temp. (degs. Cent.)

Multiply	*by*	*to obtain*
toises (French)	6	paris feet (pieds)
tons (long)	5.734×10^5	drams (avdp)
tons (long)	2.613×10^5	drams (troy)
tons (long)	1.568×10^7	grains
tons (long)	1.016×10^6	grams
tons (long)	1016	kilograms
tons (long)	3.584×10^4	ounces (avdp)
tons (long)	3.267×10^4	ounces (troy)
tons (long)	2240	pounds (avdp)
tons (long)	2722.2	pounds (troy)
tons (long)	1.12	tons (short)
Tons (metric) (T)	1000	kilograms
Tons (metric) (T)	2204.6	pounds
Tons (metric) (T)	1.1025	tons (short)
tons (short)	5.120×10^5	drams (avdp)
tons (short)	2.334×10^5	drams (troy)
tons (short)	1.4×10^7	grains
tons (short)	9.072×10^5	grams
tons (short)	907.2	kilograms
tons (short)	32,000	ounces (avdp)
tons (short)	29,166.66	ounces (troy)
tons (short)	2000	pounds (avdp)
tons (short)	2.430.56	pounds (troy)
tons (short)	0.89287	tons (long)
tons (short)	0.9078	Tons (metric) (T)
tons (short)/sq ft	9765	kg/sq meter
tons (short)/sq ft	13.89	pounds/sq inch
tons (short)/sq in	1.406×10^6	kg/sq meter
tons (short)/sq in	2000	pounds/sq inch
tons/sq mile	3.125	pounds/acre
tons/sq mile	0.07174	pounds/1000 sq ft
tons/sq mile	0.3503	grams/sq meter
tons/sq mile	350.3	kilograms/sq km
tons/sq mile	350.3	milligrams/sq meter
tons/sq mile	0.03503	milligrams/sq cm
tons/sq mile	0.03254	grams/sq ft
tons of water/24 hours	83.333	pounds of water/hr
tons of water/24 hours	0.16643	gallons/min
tons of water/24 hours	1.3349	cu ft/hr
torr (mm Hg, 0°C)	133.322	newton/meter^2
townships (U.S.)	23040	acres
townships (U.S.)	36	square miles
tuns	252	gallons
volts (abs)	10^8	abvolts

Multiply	*by*	*to obtain*
volts (abs)	3.336×10^{-3}	statvolts
volts (international of 1948)	1.00033	volts (abs)
volt/inch	.39370	volt/cm
watts (abs)	3.41304	BTU (mean)/hour
watts (abs)	0.0569	BTU (mean)/min
watts (abs)	0.01433	calories, kilogram (mean)/minute
watts (abs)	10^7	ergs/second
watts (abs)	44.26	foot-pounds/minute
watts (abs)	0.7376	foot-pounds/second
watts (abs)	0.0013405	horsepower (electrical)
watts (abs)	1.360×10^{-3}	horsepower (metric)
watts (abs)	1	joules/sec
watts (abs)	0.10197	kilogram-meters/second
watts (abs)	10^{-3}	kilowatts
watt-hours	3.415	British Thermal Units
watt-hours	3.60×10^{10}	ergs
watt-hours	2655	foot-pounds
watt-hours	859.85	gram-calories
watt-hours	1.34×10^{-3}	horsepower-hours
watt-hours	3.6×10^3	joule
watt-hours	0.8605	kilogram-calories
watt-hours	367.1	kilogram-meters
watt-hours	10^{-3}	kilowatt-hours
watt (international)	1.0002	watt (absolute)
watt/(cm^2)(°C/cm)	693.6	BTU/(hr)(ft^2)(°F/in)
wave length of the red line of cadmium	6.43847×10^{-7}	meters
webers	10^3	electromagnetic cgs units
webers	3.336×10^{-3}	electrostatic cgs units
webers	10^5	kilolines
webers	10^8	lines
webers	10^8	maxwells
webers	3.336×10^{-3}	statwebers
webers/sq in	1.550×10^7	gausses
webers/sq in	10^8	lines/sq in
webers/sq in	0.1550	webers/sq cm
webers/sq in	1,550	webers/sq meter
webers/sq meter	10^4	gausses
webers/sq meter	6.452×10^4	lines/sq in
webers/sq meter	10^{-4}	webers/sq cm
webers/sq meter	6.452×10^{-4}	webers/sq in
weeks	168	hours
weeks	10,080	minutes

Multiply	*by*	*to obtain*
weeks	604,800	seconds
yards	91.44	centimeters
yards	3	feet
yards	36	inches
yards	9.144×10^{-4}	kilometers
yards	0.91440	meters
yards	4.934×10^{-4}	miles (naut.)
yards	5.682×10^{-4}	miles (stat.)
yards	914.4	millimeters
years (sidereal)	365.2564	days (mean solar)
years (sidereal)	366.2564	days (sidereal)
years (tropical, mean solar)	365.2422	days (mean solar)
years (common)	8760	hours
years (tropical, mean solar)	8765.8128	hours (mean solar)
years (leap)	366	days
years (leap)	8784	hours
years (tropical, mean solar)	3.155693×10^{7}	seconds (mean solar)
years (tropical, mean solar)	1.00273780	years (sidereal)

2. BASIC AND SUPPLEMENTARY UNITS

A *meter (m)* is 1,650,763.73 wavelengths in vacuo of the radiation corresponding to the transition between the energy levels $2p_{10}$ and $5d_5$ of the krypton 86 atom.

A *kilogram (kg)* is the mass of the international prototype in the custody of the Bureau International des Poids et Mesures at Sevres in France.

A *second (sec)* is the interval occupied by 9,192,631,770 cycles of the radiation corresponding to the transition of the cesium-133 atom when unperturbed by exterior fields.

An *ampere* is the constant current that if maintained in two parallel rectilinear conductors of infinite length of negligible circular cross section and placed at a distance of one meter apart in vacuo would produce between these conductors a force equal to 2×10^{-7} newton per meter length.

A *kelvin* ($^\circ K$) is the degree interval of the thermodynamic scale on which the temperature of the triple point of water is 273.16 degrees.

A *candle* is such that the luminance of a full radiator at the temperature of solidification of platinum is 60 units of luminous intensity per square centimeter.

A *mole (mol)* is the amount of substance which contains as many elementary units as there are atoms in 0.012 kg of carbon-12. The elementary unit must be specified and may be an atom, an ion, an electron, a photon, etc., or a given group of such entities.

A *radian* is the angle subtended at the center of a circle by an arc of the circle equal in length to the radius of the circle.

A *steradian* is the solid angle that, having its vertex at the center of a sphere, cuts off an area of the surface of the sphere equal to that of a square with sides of length equal to the radius of the sphere.

3. DERIVED UNITS AND QUANTITIES

The *liter* was defined in 1901 as the volume of 1 kilogram of pure water at normal atmospheric pressure and maximum density equal therefore to 1.000028 dm^3. This 1901 definition applied for the purpose of the 1963 Weights and Measures Acts.

By a resolution of the 12th Conference General des Poids et Mesures (CGPM) in 1964 the word *liter* is now recognized as a special name for the dm^3, but is not used to express high precision measurements. It is used widely in engineering and the retail business, where the discrepancy of 28 parts in 1 million is of negligible significance.

A *newton (N)* is the force that, when applied to a body of mass of one kilogram, gives it an acceleration of one meter per second per second.

Stress is defined as the resultant internal force per unit area resisting change in the shape or size of a body acted on by external forces, and is therefore measured in *newtons per square meter* (N/m^2).

A *bar* is a pressure equivalent to 100,000 newtons acting on an area of one square metor.

A *joule (J)* is the work done when the point of application of a force of one newton is displaced through a distance of one meter in the direction of the force.

A *watt* is equal to one joule per second.

Dynamic viscosity is the property of a fluid whereby it tends to resist relative motion within itself. It is the shear stress, i.e., the tangential force on unit area, between two infinite horizontal planes at unit distance apart, one of which is fixed while the other moves with unit velocity. In other words, it is the shear stress divided by the velocity gradient, i.e., $(N/m^2) \div (m/sec/m) = N\ sec/m^2$.

Kinematic viscosity is the dynamic viscosity of a fluid divided by its density, i.e., $(N\ sec/m^2)/(kg/m^3) = m^2/sec$.

Density of heat flow rate (or heat flux) is the heat flow rate (W) per unit area, i.e., W/m^2.

Coefficient of heat transfer is the heat flow rate (W) per unit area per unit temperature difference, i.e., $W/m^{2\circ}C$.

Thermal conductivity is the quantity of heat that will be conducted in unit time through unit area of a slab of material of unit thickness with a unit difference of temperature between the faces; in other words, the heat flow rate (W) per unit area per unit temperature gradient, i.e., $W/[m^2(°C/m)] = W/m°C$.

The *heat capacity* of a substance is the quantity of heat gained or lost by the substance per unit temperature change, i.e., J/°C.

Specific heat capacity is the heat capacity per unit mass of the substance, i.e., J/kg°C.

Internal energy is the kinetic energy possessed by the molecules of a substance due to temperature and is measured in joules (J).

Specific internal energy (u) is the internal energy per unit mass of the substance, i.e., J/kg. When a small amount of heat is added at constant volume the increase in specific internal

energy is given by: $\mathrm{du} = c_\mathrm{v}\, dT$, where c_v is the specific heat capacity at constant volume, and dT is the increase in absolute temperature.

Specific enthalpy (h) is defined by the equation: $h = u + pv$, where p is the pressure and v is the specific volume. Specific enthalpy is measured in J/kg. When a small amount of heat is added to a substance at constant pressure, the increase in specific enthalpy is given by: $-dh = cp\, dT$, where cp is the specific heat capacity at constant pressure.

The *specific latent heat* of a substance is the heat gained per unit mass without an accompanying rise in temperature during a change of state at constant pressure. It is measured in J/kg.

The *entropy* (S) of a substance is such that when a small amount of heat is added, the increase in entropy is equal to the quantity of heat added (dQ) divided by the absolute temperature (T) at which the heat is absorbed; i.e., $dS = dQ/T$, measured in J/°K.

The *specific entropy* (s) of a substance is the entropy per unit mass, i.e., J/kg°K.

A *volt* is the difference of electric potential between two points of a conductor carrying a constant current of one ampere when the power dissipated is one watt.

A *weber* (Wb) is the magnetic flux through a conductor with a resistance of one ohm when reversal of the direction of the magnetic flux causes the transfer of one coulomb in the conductor loop.

Tesla: The magnetic flux density is the normal magnetic flux per unit area and is measured in *teslas*.

A *lumen*, the unit of luminous flux, is the flux emitted within unit solid angle of one steradian by a point source having a uniform intensity of one candle.

A *lux* is an illumination of one lumen per square meter.

Luminance is the luminous intensity per unit area of a source of light or of an illumination. It is measured in candles per square meter.

4. PHYSICAL CONSTANTS

Standard temperature and pressure (S.T.P.)	$= 273.15°\text{K}$ and $1.013 \times 10^5\ \text{N/m}^2$ $= 0°\text{C}$ and $1.013\ \text{bar}$ $= 0°\text{C}$ and $760\ \text{mm Hg}$
Molecular volume of ideal gas at S.T.P.	$= 22.41\text{liters/mol}$
Gas constant (R)	$= 8.314\ \text{J/mol°K}$
$^{RT}(273.15°\text{K})$	$= 2.271 \times 10^3\ \text{J/mol}$
Avogadro constant	$= 6.023 \times 10^{23}/\text{mol}$
Boltzmann constant	$= 1.3805 \times 10^{-23}\ \text{J/K}$
Faraday constant	$= 9.6487 \times 10^4\ \text{°C/mol}\ (= \text{A s/mol})$
Planck constant	$= 6.626 \times 10^{-34}\ \text{J sec}$
Stefan-Boltzman constant	$= 5.6697 \times 10^{-8}\ \text{W/m}^2\ \text{K}^4$
Ice point of water	$= 273.15°\text{K}\ (0°\text{C})$
Triple point of water	$= 273.16°\text{K}\ (0.01°\text{C})$
Speed of light	$= 2.998 \times 10^8\ \text{m/sec}$
Acceleration of gravity (standard) (Greenwich)	$= 9.80665\ \text{m/s}^2$ [take g as $= 9.81188\ \text{m/s}^2$ $9.81\ \text{m/s}^2$]
Universal constant of gravitation	$= 6.670 \times 10^{-11}\ \text{Newton m}^2/\text{kg}^2$
Mass of hydrogen atom	$= 1.6734 \times 10^{-27}\ \text{kg}$

5. PROPERTIES OF WATER

Temperature (°F)	Specific weight, γ (lb/ft^3)	Mass density, ρ ($\text{lb-sec}^2/\text{ft}^4$)	Dynamic viscosity, $\mu \times 10^5$ (lb-sec/ft^2)	Kinematic viscosity, $\nu \times 10^5$ (ft^2/sec)	Surface energy, $\sigma \times 10^3$ (lb/ft)	Vapor pressure, ρ (lb/in.^2)	Bulk modulus, $E \times 10^{-3}$ (lb/in.^2)
32	62.42	1.940	3.746	1.931	5.18	0.09	290
40	62.43	1.938	3.229	1.664	5.14	0.12	295
50	62.41	1.936	2.735	1.410	5.09	0.18	300
60	62.37	1.934	2.359	1.217	5.04	0.26	312
70	62.30	1.931	2.050	1.059	5.00	0.36	320
80	62.22	1.927	1.799	0.930	4.92	0.51	323
90	62.11	1.923	1.595	0.826	4.86	0.70	326
100	62.00	1.918	1.424	0.739	4.80	0.95	329
110	61.86	1.913	1.284	0.667	4.73	1.24	331
120	61.71	1.908	1.168	0.609	4.65	1.69	333
130	61.55	1.902	1.069	0.558	4.60	2.22	332
140	61.38	1.896	0.981	0.514	4.54	2.89	330
150	61.20	1.890	0.905	0.476	4.47	3.72	328
160	61.00	1.896	0.838	0.442	4.41	4.74	326
170	60.80	1.890	0.780	0.413	4.33	5.99	322
180	60.58	1.883	0.726	0.385	4.26	7.51	318
190	60.36	1.876	0.678	0.362	4.19	9.34	313
200	60.12	1.868	0.637	0.341	4.12	11.52	308
212	59.83	1.860	0.593	0.319	4.04	14.7	300

6. PERIODIC TABLE OF THE ELEMENTS (COMPLIMENTS OF THE LENOX INSTITUTE OF WATER TECHNOLOGY)

Groups ⇨ / Periods & sub-shells ⇩	I IA	2 IIA	3 IIIB	4 IVB	5 VB	6 VIB	7 VIIB	8	9 VIII	10	11 IB	12 IIB	13 IIIA	14 VIA	15 VA	16 VIA	17 VIIA	18 O
I IS	1 H 1.00794 Hydrogen																1 H 1.00794 Hydrogen	2 He 4.00260 Helium
2 2s2p	3 Li 6.941 Lithium	4 Be 9.01218 Beryllium											5 B 10.811 Boron	6 C 12.011 Boron	7 N 14.0067 Nitrogen	8 O 15.9994 Oxygen	9 F 18.9984 Fluorine	10 Ne 20.179 Neon
3 3s3p	11 Na 22.9897 Sodium	12 Mg 24.305 Magnesium											13 AI 26.9815 Aluminum	14 Si 28.0855 Silicon	15 P 30.9738 Phosphorus	16 S 32.066 Sulfur	17 CI 35.4527 Chlorine	18 Ar 39.948 Argon
4 4s3d4p	19 K 39.098 Potassium	20 Ca 40.078 Calcium	21 Sc 44.9559 Scandium	22 Ti 47.88 Titanium	23 V 50.9415 Vanadium	24 Cr 51.996 Chromium	25 Mn 54.938 Manganese	26 Fe 55.847 Iron	27 Co 58.933 Cobalt	28 Ni 58.69 Nickel	29 Cu 63.546 Copper	30 Zn 65.39 Zinc	31 Ga 69,723 Gallium	32 Ge 69.561 Germanium	33 As 74.9216 Arsenic	34 Se 78.96 Selenium	35 Br 79.904 Bromine	36 Kr 83.80 Krypton
5 5s4d5p	37 Rb 85.468 Rubidium	38 Sr 87.62 Strontium	39 Y 88.9059 Yttrium	40 Zr 91.224 Zirconium	41 Nb 92.9064 Niobium	42 Mo 95.94 Molybdenum	43 Tc (98) Technetium	44 Ru 101.07 Ruthenium	45 Rh 102.906 Rhodium	46 Pd 106.42 Palladium	47 Ag 107.868 Silver	48 Cd 112.411 Cadmium	49 In 114.82 Indium	50 Sn 118.710 Tin	51 Sb 121.75 Antimony	52 Te 127.60 Tellurium	53 I 126.90 Iodine	54 Xe 131.29 Xenon
6 6s4f5d6p	55 Cs 132.905 Cesium	56 Ba 137.327 Barium	57 La 138.906 Lanthanum	72 Hf 178.49 Hafnium	73 Ta 180.948 Tantalum	74 W 183.85 Tungsten	75 Re 186.207 Rhenium	76 Os 190.2 Osmium	77 Ir 192.22 Iridium	78 Pt 195.08 Platinum	79 Au 196.97 Gold	80 Hg 200.59 Mercury	81 TI 204.383 Thallium	82 Pb 207.2 Lead	83 Bi 208.98 Bismuth	84 Po (209) Polonium	85 At (210) Astatine	86 Rn (222) Radon
7 7s5f6d	87 Fr (223) Francium	88 Ra (226) Radium	89 Ac (227) Actinium	104 Rf (261) Ruther-fordium	105 Ha (262) Dubnium	106 Sg (263) Seaborgium	107 Ns (262) Bohrium	108 Hs (265) Hassium	109 Mt (266) Meitnerium	110		112						

6 4f	58 Ce 140.116 Cerium	59 Pr 140.91 Praseody-mium	60 Nd 144.24 Neodymium	61 Pm (145) Promethium	62 Sm 150.35 Samarium	63 Eu 107.26 Europium	64 Gd 157.25 Gadolinium	65 Tb 158.925 Terbium	66 Dy 162.50 Dysprosium	67 Ho 104.930 Holmium	68 Er 167.26 Erbium	69 Tm 168.934 Thulium	70 Yb 173.04 Ytterbium	71 Lu 174.967 Lutetium
7 5f	90 Th 232.038 Thorium	91 Pa (231) Protactinium	92 U 238.029 Uranium	93 Np (237) Neptunium	94 Pu (244) Plutonium	95 Am (243) Americium	96 Cm (247) Curium	97 Bk (247) Berkelium	98 Cf (251) Californium	99 Es (252) Einsteinium	100 Fm (257) Fermium	101 Md (258) Mendelevium	102 No (259) Nobelium	103 Lr (262) Lawrencium

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