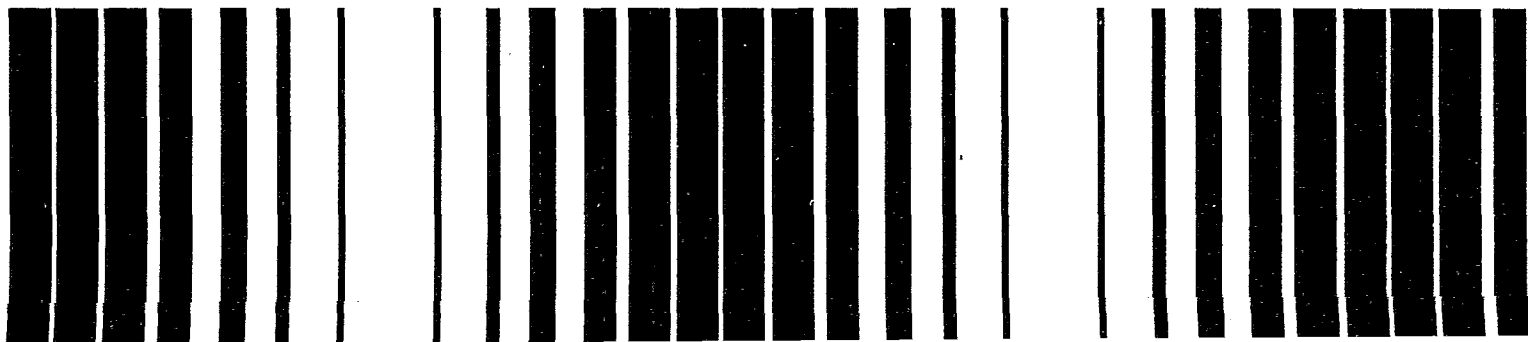**♺EPA**

Manual

Nitrogen Control

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U.S. Environmental Protection Agency

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Notice

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Contents

Contents (continued)

List of Figures

Figures (continued)

List of Tables

Tables (continued)

Tables (continued)

Acknowledgments

Chapter 1
Introduction

1.1 Background and Purpose

The growing demand for water resources has generated an equivalent need for effective water and wastewater management strategies. This need is heightened by an awareness that a sound hydrosphere is fundamental to the world's ecology. In the United States these driving forces have found support in the regulations that mandate the proper handling and treatment of wastewaters discharged to the aquatic environment.

Application of processes to control wastewater discharges to the aquatic environment began around the turn of the century, generally addressing the removal of particulates and oxygen-demanding carbonaceous materials. The removal of other constituents did not receive substantive attention until the latter part of this century. The control of nitrogen has subsequently been identified as an important environmental activity, demonstrated by the adverse effects that excess levels of different forms of nitrogen have had on aquatic systems. Ammonium-nitrogen oxidation processes were fairly widely applied in England by mid-century. Such processes were not implemented on a significant scale in the United States until the late 1960s, although early activated sludge and trickling filter plants often did nitrify, even if not explicitly designed to do so. The first full-scale application of a nitrogen removal process took place in South Lake Tahoe, California, in 1969, although, as a relatively new technology, it experienced problems.

Research and development activities through the mid-1970s served as the basis for the U.S. Environmental Protection Agency's (EPA's) first comprehensive manual of design for various nitrogen control technologies (1). The *Process Design Manual for Nitrogen Control*, published in 1975, covered a broad spectrum of processes, reflecting the diverse approaches being evaluated and applied at the time. Its intent was to present design information for technologies that appeared to have a viable, practical application to nitrogen control. Such processes can be divided to two broad categories. The first group provides for the conversion of organic and ammonium nitrogen by oxidation to nitrate nitrogen. These are biological processes and are generally termed nitrification.

The second group accomplishes the actual removal of nitrogen from the wastewater, not simply the conversion from one form to another. This is also typically done by biological means, coupling an anoxic denitrification step with nitrification. Physical/chemical processes were also presented for nitrogen removal, including ion exchange, ammonia stripping, and breakpoint chlorination.

Since the first manual's publication, the trend in nitrogen control technology applications has been overwhelmingly in favor of biological processes, with only a few instances in which physical/chemical processes have been implemented. Virtually all of the greater than 15,000 publicly owned treatment works (POTWs) in the United States use biological processes to remove organics. While the total number of facilities is projected to rise by approximately 30 percent to meet the demands of the full U.S. population, those systems identified specifically for biological nitrogen control are expected to at least double in number (2,3). Biological processes are proven and well demonstrated and lend themselves most efficiently to expansion or upgrade for biological nitrification or total nitrogen removal. Biological processes also minimize the use of often expensive and sometimes environmentally incompatible chemicals and will generally achieve residual ammonium levels that are lower than can be effectively accomplished by most physical/chemical processes.

Breakpoint chlorination, ion exchange, and air stripping processes received considerable attention in the 1975 manual, but have been employed in a very limited number of POTW applications over the last 15 years. This is not to say that these technologies are not viable. Practical applications for these nonbiological controls exist in some POTW settings, such as very cold climates or for industrial pretreatment. Where particularly high concentrations of ammonium nitrogen exist (e.g., landfill leachates), both physical/chemical and biological processes are utilized. Conversely, significant attention is being given to natural processes for nitrogen control, particularly for application to small systems. These configurations, including natural and constructed marshes and wetlands are generally considered to be in an emerging, developmental state at

this point, although with potentially significant, widespread applications.

This manual is an update and revision of the original 1975 edition. It strives to maintain the high technical quality and generous provision of reference materials provided by the 1975 edition, although it also represents a significant shift in overall content. Given the experience of the past 18 years, the focus of this second edition is directed to those biological/mechanical systems that have found widespread use for nitrification and nitrogen removal. Design criteria for physical/chemical systems are not provided; however, there is a brief discussion of such processes in Chapter 2, in which their applicability under specific site conditions and wastewater applications is addressed. The design of natural systems also is not considered within the context of this manual, except in the planning and development of alternatives for technology selection, a point of discussion within Chapter 2. Adequate references are given in Chapter 2 to assist the reader in seeking design information on both natural systems and physical/chemical processes.

The primary audience is the designer of small- to medium-sized facilities, although the application of the manual is not limited to any range of plant sizes. The intent is to provide a manual that can be readily used by those who do not necessarily specialize in the design of nitrogen control facilities. Detailed theoretical discussions are not provided. Rather, the manual focuses on the major process and design aspects considered in the development of an effective design. It begins with process basics and proceeds to the presentation of detailed design criteria and the development of process designs, using examples to demonstrate calculation sequences. In addition, the manual is organized to help the designer in the planning stages of a facility, highlighting important process and operation and maintenance (O&M) considerations. The intention is to give direction before plans advance to a level at which redesign would be costly. For design purposes this manual is most prudently used in its entirety.

The manual also is assembled for use as a desk reference, or handbook. The table of contents is detailed enough for the user to be able to find specific topics quickly, and an effort has been made to present information as often as possible in the form of charts, tables, and figures. In addition to aiding designers, an objective of the manual is to serve as a source for reviewers, operators, regulators, and manufacturers.

1.2 Scope of the Manual

The manual is constructed to progress from a broad discussion of nitrogen in the environment, to the concepts of using biological processes to control or remove nitrogen, and finally to the details of designing specific systems. This first chapter describes the relationships of nitrogen in the environment. The fundamental purpose of the manual, implementation of nitrogen controls in municipal systems, is brought into focus in Chapter 2 by outlining design principles. An array of issues are presented that enter into the designer's strategy. The chapter discusses the relative importance of each issue in order to help the designer avoid pitfalls, before they are compounded by the growing detail of design. Chapters 3 and 4 give the theoretical foundations of the nitrification and denitrification processes by drawing on concepts of microbiology, chemistry, and kinetics. Although the discussions of mathematical and computer modeling in Chapter 5 are intended to stand independently from the other chapters, they provide a useful bridge between Chapters 3 and 4 and Chapters 6, 7, and 8.

In the latter three chapters, the conceptual bases of nitrification and denitrification are developed to design criteria, and design examples are presented to assist in producing a specific configuration that will meet performance objectives. Chapter 6 addresses suspended growth and attached growth configurations for biological nitrification. Denitrification processes are addressed in Chapter 7, but only as applied in a separate stage using a supplemental carbon source. The current trend has been to accomplish nitrification and denitrification in single-sludge systems, using wastewater carbon for the denitrification step. Single-sludge systems, which are addressed in Chapter 8, have seen increasing application in lieu of the alternative two- or three-sludge systems for nitrogen removal.

A significant portion of the information in this manual has been carried forward from the 1975 document. Substantial modifications have been made, however, in directing the emphasis toward biological/mechanical systems. Contributions to this work were developed from the following sources: 1) the experience of the individuals involved in the preparation of the manual; 2) the EPA research, development, and demonstration program; 3) existing literature; 4) design/performance and operating experience at existing nitrogen control facilities; 5) progress reports regarding ongoing projects; 6) private communication with investigators active in the field; and from 7) operating personnel at existing wastewater treatment plants.

The material presented is a distillation of knowledge reflecting relatively few generations of full-scale experience with nitrogen control technologies. As such, it represents a perspective of the present state-of-the-art, and not necessarily a complete understanding of the technology. Enhanced understanding and knowledgeable application of current demonstrated approaches, coupled with new technologies or approaches to join those that are emerging, represent the expectation and challenge of the future.

1.3 How to Use This Manual

A perusal of the Table of Contents gives the reader an overview of the subject matter contained in this manual. All users should read Chapter 2, which serves to set the background and protocol for effective design. It is a lead-in to the design chapters (6, 7, and 8) and establishes the design examples used in these chapters. The following chapter-by-chapter description provides a more detailed presentation of the contents and the objectives of each chapter.

Chapter 1, *Introduction,* first gives the reader a broad view of the manual, a perspective on how designers can use the manual to serve their needs, and an explanation of the manual's limitations. The remaining sections give a general overview of the cause-and-effect relationships of nitrogen in the environment and why nitrogen removal is important. Sources of nitrogen are identified and the movement and transformation of nitrogen's various forms are presented. The effects of nitrogen within the aquatic environment are then discussed, giving the designer a perspective on how a treated wastewater discharge affects the local environment. From this, a brief discussion is presented of the regulatory structure that is in place to evaluate and set discharge limits with respect to nitrogen (and other constituents) and to ensure the quality and beneficial use of the receiving water.

Chapter 2, *Total System Design with Nitrogen Control,* presents the groundwork necessary to initiate the design process. An overview of technology concepts, process selection considerations, and special issues that require a multidimensional approach are discussed. Physio-chemical and natural systems approaches are treated from the perspective of appropriate technology selection. Chapter 2 also introduces the design examples, which provide a common basis for illustrative designs developed in Chapters 6 through 8.

Chapter 2 is an important element of the manual as a whole and should be read by all who intend to use the manual in support of a facility design or review. The chapter is designed to give an overview to the considerations and calculations that are addressed in the early stages of facility design and offers mass-balancing methods and common design information from which checks can be made of an ongoing design process. The chapter addresses key considerations when defining site requirements and the bases for selecting alternative technologies. Careful attention to such issues during the early stages of a project will minimize the need for redirection or redesign and help to avoid the common pitfalls encountered in the design of a facility.

Chapter 3, *Process Chemistry and Kinetics of Biological Nitrification,* offers the designer the fundamentals of the microbiology of nitrifying organisms, their means of metabolism, the biochemical pathways these organisms use

to nitrify, and the stoichiometry useful in materials balancing. The relevant kinetic rate expressions are given, focusing on the intrinsic and controlling environmental factors associated with the process. Chapter 3 serves as the basis for design concepts and criteria presented in Chapters 6 and portions of 8.

Chapter 4, *Process Chemistry and Kinetics of Biological Denitrification,* presents the microbiology, the metabolism, and the stoichiometry of denitrifying organisms in a format similar to that used in Chapter 3. Rate expressions are also similar, but the specifics of the controlling environmental factors, such as pH, dissolved oxygen (DO), temperature, and inhibitors, differ substantially. Effective use and understanding of Chapters 7 and 8 follow from the basic factors given in Chapter 4.

Chapter 5, *Mathematical Modeling of Nitrification and Denitrification,* establishes the conceptual and mathematical frameworks that integrate the theoretical concepts of Chapters 3 and 4 with quantified design experience. The objective is to make the designer aware of how models can serve as effective tools for formulating a design, while at the same time warning of the hazards of using such models without a detailed understanding of the processes and an adequate data base. This chapter outlines the phases of model development and the various uses of models. A number of different types of models are presented and one, the IAWPRC model, is analyzed as an example. This chapter need not be read to understand the design material in Chapters 6, 7, and 8.

Chapter 6, *Design Considerations for Biological Nitrification Processes,* is a sequel to Chapter 3. Topics relevant to the development of design criteria for nitrification systems are divided between those applicable to suspended growth systems and those applicable to attached growth systems. The suspended growth section emphasizes the application of kinetic theory and highlights both solids residence time and specific rate approaches. Designs for various completely mixed and plug flow configurations are analyzed. The attached growth design criteria are developed more from empirical approaches and less from kinetic theory, drawing from an accumulation of design experience. Trickling filters, rotating biological contactors, and other configurations of attached growth systems are presented. The objective of this chapter is to direct the designer in the specifics of selecting, linking, and sizing the various unit operations for nitrification.

Chapter 7, *Design Considerations for Denitrification Processes with Supplemental Substrate Addition.* Today, separate-stage denitrification processes with a supplemental substrate are generally applied using suspended growth, packed bed, or fluidized bed systems. Design considerations and case studies are presented for each of these three system configurations. The requirements surrounding the handling, storage, feed, and control of methanol are also presented, since methanol is essen-

tially the only supplemental substrate used in current separate-stage denitrification systems. The treatment for removal of excess methanol, if needed, by post-aerobic treatment is also discussed in Chapter 7.

Chapter 8, *Design Considerations for Single-Sludge Nitrification-Denitrification Processes*, classifies the various single-sludge nitrogen removal technologies into five categories according to their number of stages, phases, and anoxic zones. The more common proprietary and nonproprietary single-sludge process configurations for nitrogen removal are reviewed by comparing their design criteria and expected process performance against their unique design features. Process design scenarios are described and a sample configuration for each scenario is selected from among the classifications. These representative examples demonstrate controlling design factors, such as recycle and aeration.

1.4 Nitrogen in the Environment

The total mass of Earth's nitrogen circulates within the biosphere among four main banks: the atmosphere, the hydrosphere, the Earth's crust, and the tissues of living and dead organisms. Each bank contains inventories of nitrogen in its various forms. Although the total is unchanging, the inventories of the various forms are in constant flux. In a natural world, relative to one influenced by the activities of people, this flux could be considered approximately steady-state. However, there are two activities that cause significant non-steady-state shifts in the baseline nitrogen balance: 1) the mining and use of nitrogen-containing minerals and fossil fuels that have long been out of active nitrogen circulation and 2) a positive net fixation of nitrogen gas attributable to chemical manufacturing and the cultivation of leguminous, nitrogen-fixing plants. Unfortunately, the hydrosphere has become a sink for much of the excess nitrogen that has been mobilized as a result of human commerce. Consequently, some significant undesirable effects result from the accumulation of nitrogen within niches of the aquatic environment. The quality of water may be affected and the potential beneficial uses of natural water systems may be diminished. Nitrogen, in its various forms, can deplete DO levels in receiving waters, stimulate aquatic growth, exhibit toxicity toward aquatic life, present a public health hazard, and affect the suitability of wastewater for reuse.

This manual presents the design of processes and technological configurations that will control and/or remove nitrogen from wastewaters discharged to natural waters. It is appropriate, therefore, to first review the nature of nitrogen and its presence in the environment. The remainder of this chapter 1) reviews how the various nitrogen forms are physically transported and chemically changed in what is known as the nitrogen cycle (Section 1.4.1); 2) surveys the major sources of both natural and human-caused nitrogen loading (Section 1.4.2); and 3) discusses some of the effects that nitrogen compounds have in altering the environment (Section 1.5). Finally, a closing section (Section 1.6) presents a brief discussion of the regulatory structure by which nitrogen control requirements and performance levels are implemented for wastewater treatment plants in the United States.

1.4.1 The Nitrogen Cycle

Nitrogen exists in many forms in the environment. The movement and transformation of these nitrogen compounds through the biosphere is characterized by the nitrogen cycle, a general depiction of which is shown in Figure 1-1. The atmosphere serves as a reservoir of nitrogen in the form of nitrogen gas. Although virtually inexhaustible (the atmosphere is 79 percent nitrogen), the nitrogen must be combined with hydrogen or oxygen before it can be assimilated by higher plants; the plants, in turn, are consumed by animals. Human intervention through industrial nitrogen fixation processes and the large-scale cultivation of nitrogen-fixing legumes has played a significant role in altering the historical nitrogen cycle. The amount of nitrogen fixed annually by these two mechanisms now exceeds by as much as 10 percent the amount of nitrogen fixed by terrestrial ecosystems before the advent of agriculture (4).

Nitrogen can form a variety of compounds because of the different oxidation states it can assume. In the environment, most changes from one oxidation state to another are brought about biologically. Consider the nitrogen forms that are of interest in the soil/water environment:

Nitrogen Compound	Formula	Oxidation State
Ammonia	NH_3	−3
Ammonium ion	NH_4^+	−3
Nitrogen gas	N_2	0
Nitrite ion	NO_2^-	+3
Nitrate ion	NO_3^-	+5

The un-ionized, molecular ammonia exists in equilibrium with the ammonium ion, the distribution of which is dependent upon system pH and temperature; in fact, very little ammonia exists at pH levels less than neutral. This is an important relationship and is discussed in greater detail in Section 1.5.3 and in Chapters 3 and 6.

Transformation of these nitrogen compounds can occur through several mechanisms. Those of importance include fixation, ammonification, synthesis, nitrification, and denitrification. Each can be carried out by particular microorganisms with either a net gain or loss of energy; energy considerations often play an important role in determining which reactions occur.
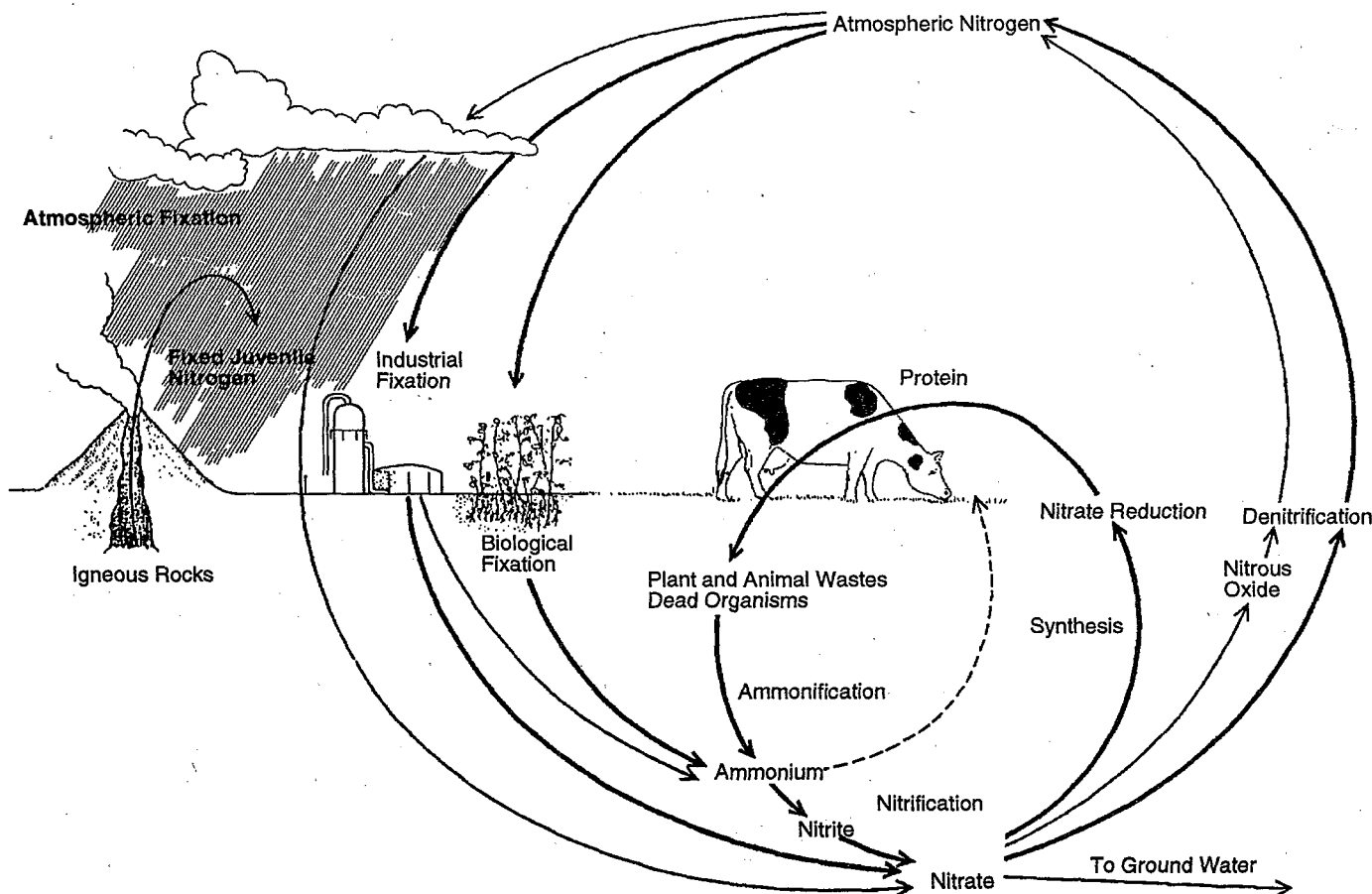
Figure 1-1. The nitrogen cycle.

Fixation of nitrogen means the incorporation of inert, gaseous nitrogen into a chemical compound such that it can be used by plants and animals. Fixation of nitrogen from N_2 gas to organic nitrogen is predominantly accomplished biologically by specialized microorganisms and the associations between such microorganisms and plants (5). Atmospheric fixation by lightning and industrial fixation processes (fertilizer and other chemicals) plays a smaller, but significant, role as a fixation method.

Fixation Process		Product
	→ biological	→ organic nitrogen compounds
N_2 gas	→ lightning	→ nitrate
	→ industrial	→ ammonium, nitrate

Ammonification is the change from organic nitrogen to the ammonium form. An important hydrolysis reaction involves urea, a nitrogen compound found in urine:

$$H_2NCONH_2 + 2H_2O \xrightarrow[\text{urease}]{\text{enzyme}} (NH_4)_2\ CO_3$$

urea   ammonium carbonate
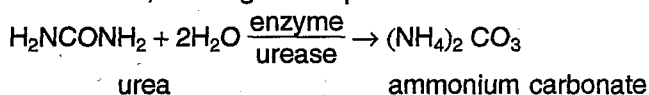
In general, ammonification occurs during decomposition of animal and plant tissue and animal fecal matter:

organic nitrogen + microorganisms → NH_3/NH_4^+

    (protein, amino acids, etc.)
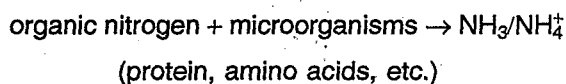
Synthesis, or assimilation, is a biochemical mechanism that uses ammonium or nitrate compounds to form plant protein and other nitrogen-containing compounds:

$NO_3^- + CO_2$ + green plants + sunlight → protein

$NH_3/NH_4^+ + CO_2$ + green plants + sunlight → protein

Animals require protein from plants and other animals. With certain exceptions, they are not capable of transforming inorganic nitrogen into an organic nitrogen form.

Nitrification is the biological oxidation of ammonium. This is done in two steps, first to the nitrite form, then to the nitrate form. Two specific chemoautotrophic bacterial genera are involved, using inorganic carbon as their source of cellular carbon:

$$NH_4^+ + O_2 \xrightarrow[\text{bacteria}]{\text{Nitrosomonas}} NO_2^- + O_2 \xrightarrow[\text{bacteria}]{\text{Nitrobacter}} NO_3^-$$

ammonium   nitrite   nitrate
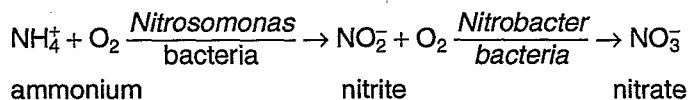
The transformation reactions are generally coupled and proceed rapidly to the nitrate form; nitrite levels at a given time are relatively low. The nitrate formed may be used in synthesis to promote plant growth, or it may be subsequently reduced by denitrification, as suggested by Figure 1-1.

Denitrification is the biological reduction of nitrate to nitrogen gas. It can proceed through several steps in the biochemical pathway, with the ultimate production of nitrogen gas. A fairly broad range of heterotrophic bacteria are involved in the process, requiring an organic carbon source for energy:

$$NO_3^- + \text{organic} \rightarrow NO_2^- + \text{organic} \rightarrow N_2 + \text{carbon} + \text{water}$$

carbon	carbon	dioxide
nitrate	nitrite	nitrogen
		gas
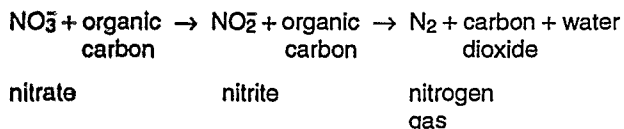
It is important to note that if both oxygen and nitrate are present the bacteria will typically preferentially use oxygen in the oxidation of the organic matter because it yields more energy. Thus, for denitrification to proceed, anoxic conditions must usually exist, although this is not strictly the case for all bacteria.

The ammonification, synthesis, nitrification, and denitrification reactions are the primary mechanisms employed in the treatment of wastewaters for nitrogen control and/or removal. Detailed discussions of their microbiological, chemical, and kinetic aspects may be found in Chapters 3 (nitrification) and 4 (denitrification).

The transport mechanisms primarily responsible for the movement of nitrogen through the environment are precipitation, dustfall, sedimentation in water systems, wind, ground-water movement, stream flow, overland runoff, and volatilization. Although these are not mechanisms by which transformations take place, they can cause a change in the environment whereby conditions will force change and transformations will occur. Environmental conditions that affect the behavior of reactions include temperature, pH, microbiology, oxidation/reduction potential, and the availability of substrate, nutrients, and oxygen. Although transport and transformation mechanisms are described as individual processes, it is useful to understand that these comprise a dynamic continuum and there may be no distinct boundary governing the transformation of specific forms of nitrogen.

Since this manual's purpose is to address wastewater treatment for nitrogen control, it is of interest to differentiate between the surface water and sediment environment and the soil–ground-water environment of the nitrogen cycle, which are directly affected by wastewater treatment practice. This aids in understanding the roles that nitrogenous compounds play in each and the problems that can be encountered.

1.4.1.1 The Nitrogen Cycle in Surface Waters and Sediments

A representation of the nitrogen cycle that is applicable to the surface water environment is presented in Figure 1-2 (6). As shown, nitrogen can be added by atmospheric deposition through precipitation and dustfall; surface run-off; subsurface ground-water entry; and direct discharge of wastewater effluent. In addition, nitrogen gas from the atmosphere can be fixed by certain photosynthetic blue-green algae and some bacterial species.

Ammonification, nitrification, synthesis, and denitrification can occur within the aquatic environment. Ammonification of organic matter is carried out by microorganisms. The ammonium thus formed, along with nitrate, can be assimilated by algae and aquatic plants for synthesis. If excessive, such growths may create water quality problems.

Biological nitrification of ammonium can occur; if significant, it can result in depletion of the DO content of the water (the oxidation of 1.0 mg/L of ammonium-nitrogen will consume 4.6 mg/L of oxygen). Denitrification produces nitrogen gas, which may escape to the atmosphere. Because anoxic conditions are required for denitrification, the oxygen-deficient hypolimnion (or lower layer) of lakes and the sediment zone of streams and lakes are important zones of denitrification activity (7).

1.4.1.2 The Nitrogen Cycle in Soil and Ground Water

Figure 1-3 shows the major aspects of the nitrogen cycle associated with the soil–ground-water environment (8). Nitrogen can enter the soil from the application of wastewater or sewage treatment plant effluent, artificial fertilizers, plant and animal matter, precipitation, and dustfall. In addition, nitrogen-fixing bacteria in the soil convert nitrogen gas into forms available to plant life. Humans have increased the amount of nitrogen fixed biologically by cultivation of leguminous crops (e.g., peas and beans).

Usually more than 90 percent of the nitrogen present in soil is organic, either in living plants and animals or in humus originating from decomposition of plant and animal residues. Nitrate content is generally low because it is taken up for synthesis, it is leached by water percolating through the soil, and because of denitrification activity below the aerobic top layer of soil. Synthesis and denitrification rarely remove all nitrates added to the soil from fertilizers and nitrified wastewater effluents. As such, elevated ground-water nitrates leached from soil are a major ground-water quality problem in many areas.

1.4.2 Sources of Nitrogen

An understanding of the various sources of nitrogenous materials offers a perspective on the relative contribution that municipal wastewater treatment plants make to the cumulative effects of nitrogen in the environment. This, in turn, can influence decisions regarding the level and type of treatment that is most suited for a particular location. In analyzing a nitrogen pollution problem, care must be taken to ensure that all possible sources of pollution are investigated and that the amount to be expected from each is accurately estimated. Once estimates are made and used in concert with water quality studies, nitrogen
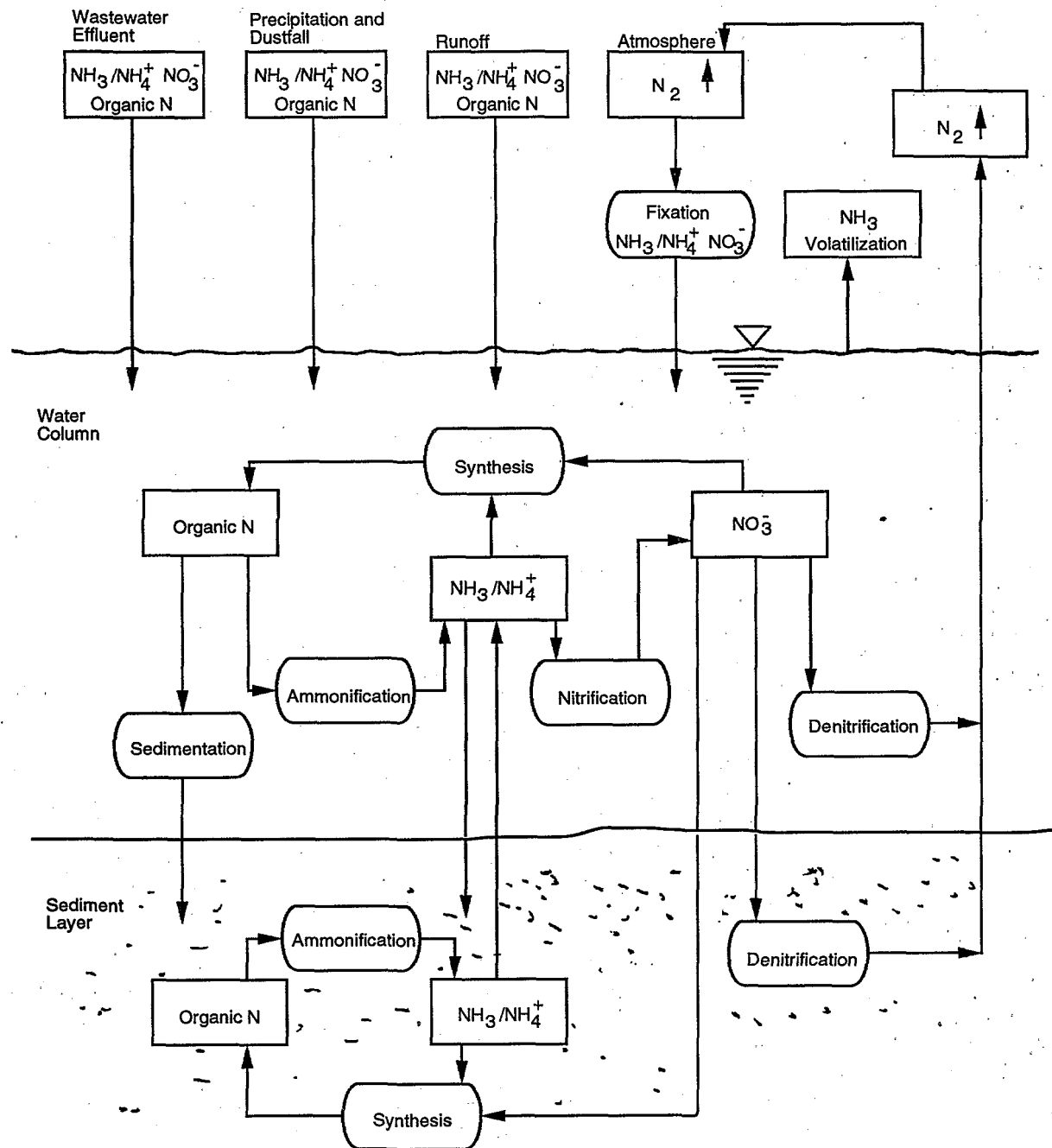
Figure 1-2. The nitrogen cycle in surface water.
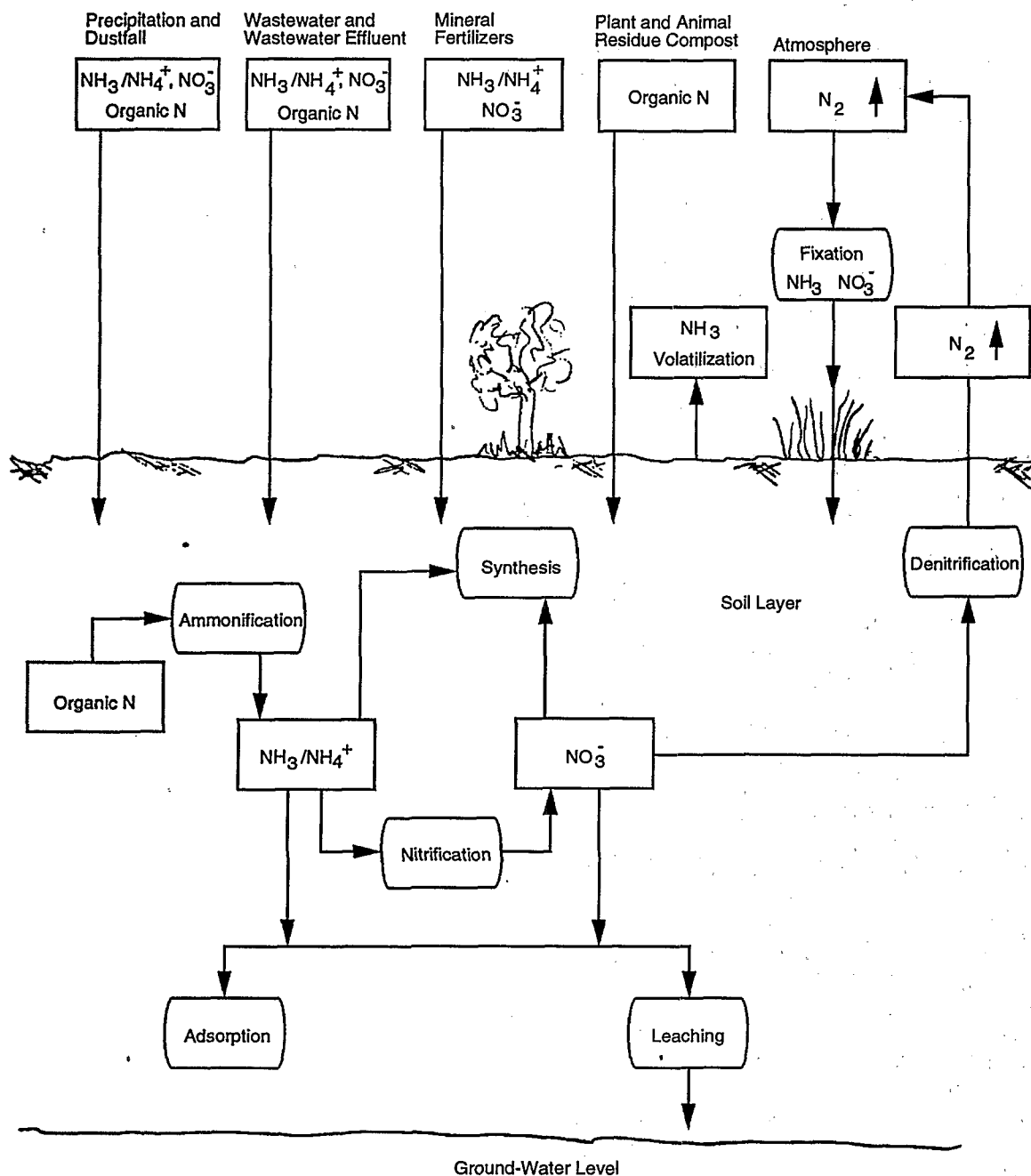
Figure 1-3. The nitrogen cycle in soil and ground water.

control measures can be oriented to the more significant sources, and specifically to the type of impacts unique to the individual sources. As an example, an estuary with excessive aquatic growth may receive nitrogen-containing effluent from a sewage treatment plant, urban runoff, and runoff from animal feedlots. The nitrogen from the treatment plant is shown to be a small fraction of the contribution from the other two sources, yet further study of the geometry and flushing action of the estuary shows significant local toxicity impacts from the plant's discharge. Since the treatment plant plays no significant role in the total nitrogen load to the estuary and its consequent biostimulatory effects, it should not be required to remove nitrogen from its discharge. However, nitrification of the plant discharge may be needed to eliminate localized ammonia toxicity effects.

Nitrogenous materials can enter the aquatic environment from either natural or human-caused sources. The proper delineation of these sources can often be clouded, since the apparent quantities from natural sources can include nitrogen generated from human activity. For example, while nitrogen fixed by lightning may be expected in rainfall, the combustion of fossil fuels or the application of liquid ammonia agricultural fertilizers, with subsequent release to the air through volatilization, can increase rainfall concentrations of nitrogen substantially. The pervasiveness of human impact on the environment limits the certainty with which naturally occurring nitrogen and pollution source nitrogen can be differentiated quantitatively. Measurements of naturally occurring baseline nitrogen levels are best made in more remote, undeveloped and pristine areas; yet interpretation of these data should still be considered uncertain to some degree.

The several direct sources from which the bulk of nitrogen enters the aquatic environment are listed in Table 1-1, along with the principal transport mechanisms responsible for delivering that nitrogen to natural receiving water systems. Nitrogen originates from a source in soluble and/or particulate form; it may also change form en route to the receiving waters. Soluble nitrogen is primarily in the inorganic form of ammonia, ammonium, nitrite, and nitrate. Lesser amounts of reduced soluble nitrogen are in the form of low molecular weight organic compounds such as urea and proteins. Particulate nitrogen can also be either organic or inorganic in nature. Particulate organic nitrogen can consist of insoluble synthetic compounds, flocculated one-celled microbes, and particles of detritus from animal and plant tissues and animal wastes. Inorganic particulate nitrogen typically consists of inorganic species adsorbed or ionically exchanged onto sediment that becomes suspended in water.

1.4.2.1 Natural Sources

Natural sources or transport mechanisms of nitrogen substances include atmospheric precipitation, dustfall, nonurban and nonagricultural runoff, and biological fixation.

Table 1-1. Major Direct Sources of Nitrogen

Source	Principal Transport to Natural Water Systems
Untreated sewage	Direct discharge
POTW effluent	Direct discharge, land application
POTW waste solids	Direct discharge, land application
Industrial sources	Direct discharge, ground-water movement, precipitation
Volcanoes and other earth emissions	Precipitation, wind, and gravitational settling
Fertilized farms	Surface runoff, ground-water movement
Animal wastes	Volatilization/precipitation, surface runoff, ground-water movement
Decaying plants and animal tissue	Surface runoff, ground-water movement
Septic tanks and leaching fields	Ground-water movement
Ship/other vessels	Direct discharge
Urban surfaces	Direct discharge, surface runoff
Fossil fuels	Precipitation, wind and gravitational settling
Nitrogen-fixing organisms	In situ
Disturbed earth	Wind and gravitational settling
Landfill leachate	Ground-water movement

Nitrogen measured in precipitation is most often a result of both soluble and particulate nitrogen forms scrubbed from the atmosphere. Natural components would include nitrogen oxides fixed by lightning and emitted from volcanic eruptions, wind-blown dust originating from natural areas, and ammonia released from decaying animal and plant matter. As examples, total nitrogen in rainfall in Sweden was cited as 0.2 ppm, while the average concentration of nitrogen in western U.S. snows, mainly in the Sierra Nevada Mountains, was 0.15 ppm ammonium-nitrogen, 0.01 ppm nitrite-nitrogen, and 0.02 ppm nitrate-nitrogen (5). Again, the degree to which such values are representative of "natural" conditions cannot be determined with any certainty.

The quantities of nitrogen in nonurban runoff from non-fertilized land may be expected to vary greatly, depending on the erosive characteristics of the soil and the migration of dustfall and precipitation originating from proximate urban and/or industrial areas. As an example, runoff from forested land in Washington contained 0.13 mg/L nitrate-nitrogen and 0.20 mg/L total nitrogen (7). A survey of surface runoff from 90 percent forested land in the east-

ern, central, and western United States yielded mean total nitrogen concentrations of 0.19, 0.06, and 0.07 mg/L, respectively. Concentrations for 50 percent forested lands of the same regions are approximately double these values at 0.34, 0.25, and 0.18 mg/L, respectively (9).

Biological fixation may add nitrogen to surface water directly or to ground water via leaching from soil. Of particular interest is the role of fixation in the eutrophication of lakes. Certain photosynthetic blue-green algae, such as the species of *Nostic*, *Anabaena*, *Gleotrichia*, and *Calathrix*, are common nitrogen fixers. As an example, as much as 14 percent of the total nitrogen entering eutrophic Lake Mendota in Wisconsin was added by fixation (6). Finally, lightning contributes notably to the mass quantity of fixated nitrogen. One estimate reports that approximately 15 percent of all naturally fixed nitrogen results from lightning (4).

1.4.2.2 Human-Caused Sources

Sources of nitrogen related to human activity include untreated and treated domestic sewage and industrial wastes, leachates, atmospheric deposition, and surface runoff. These general areas are presented in the following discussions.

Domestic Sewage. Untreated sewage flowing from municipal collection systems typically contains 20–85 mg/L total nitrogen. Flow from residences, a major component of domestic wastewater, has been observed to contain 30–80 mg/L TKN (total Kjeldahl nitrogen). The total nitrogen in domestic sewage comprises approximately 60 percent ammonium nitrogen, 40 percent organic nitrogen, and very small quantities of nitrates. Unit loading from residences has been estimated at an average of 0.012 kg N/capita/d (0.027 lb/capita/d). Septage from rural septic system tanks is often collected and mixed with untreated sewage at the head of treatment works. Septage nitrogen content is 100–1,600 mg/L TKN with 700 mg/L TKN being a typical value (9). The volume of septage is generally low relative to wastewater plant flow, although in smaller systems it can be a significant input. The nitrogen content of combined sewer overflows (CSO) is often calculated from the weighted averaging of storm water and domestic sewage concentrations. Nitrogen concentrations of CSO vary radically over a rainfall event, particularly as an effect of the first wash of drainage area surfaces and the first flush of settled solids in the drainage collection grid.

Treated domestic sewage will exhibit a variety of nitrogen levels, depending on the type of treatment applied. A typical activated sludge facility reduces total nitrogen content by virtue of cell synthesis and solids removal. Most ammonia, however, passes through unless there is a specific treatment provision for nitrification. Reductions in total nitrogen beyond 20–30 percent would require

denitrification. Conventional activated sludge treatment generally yields effluent nitrogen levels of 15–35 mg/L total nitrogen. Advanced biological nitrification-denitrification treatment can generally yield an effluent quality of 2–10 mg/L total nitrogen (9). Sludges generated from the treatment of wastewaters will also contain nitrogen and will serve as a nitrogen source, depending on the method of disposal. Section 2.4.1, provides a discussion of expected effluent nitrogen levels for various technology applications to municipal wastewaters.

Industrial Wastewaters. Industry adds nitrogen into aqueous waste streams as a result of the use of water in processing and the secondary use of water to scrub gaseous waste streams. Some industries that yield more elevated levels of nitrogen in their primary or process water include chemical (fertilizer and nitrogenous compound manufacturing); paper and pulp (natural products, ammonia-based pulping); mining and metals (ore processing, nitric acid pickling); and food processing (protein-enriched wash and cooking waters). As an example, estimates from Sweden suggest that these primary industrial categories account for 95 percent of the country's industrial aqueous nitrogen discharge. Textiles and leather manufacturing account for the remaining 5 percent (10). Industrial and power generation facilities use water for scrubbing flue gases, which creates secondary nitrogen-containing waste streams. Nitrogen oxides, created during combustion of fossil fuels, hydrolyze to nitrate when solubilized in water.

Landfill Leachates. Leachate from municipal solid waste landfills (MSWLs) is characterized as a relatively low-volume, high-strength wastewater. A survey of leachate characterization studies for many MSWLs shows ammonium values of 0–1,160 mg/L and nitrate plus nitrite nitrogen of 0.2–10.3 mg/L. Treatment design has to be flexible to allow for the typical high degree of variability in leachate strength observed with landfill age (11).

Atmospheric Deposition. Atmospheric nitrogen generally enters an aquatic environment in one of two forms: inorganic nitrogen that is solubilized in rainwater or particulate organic and mineralized nitrogen that either settles by gravity or is scrubbed by rain onto the receiving water surface. Atmospheric deposition over the land area of a watershed is often accounted for in surface runoff loads; therefore, atmospheric contributions to a waterbody's nitrogen loading is usually attributed only to precipitation that falls directly over the water surface. Most soluble, inorganic nitrogen originates from volatilization of ammonia-nitrogen and combustion of fossil fuels. Combustion oxidizes nitrogen contained in oil and coal and (at higher temperatures) free atmospheric nitrogen. Nitric oxide is the principal product, which rapidly oxidizes to nitrogen dioxide. Nitrous oxide can also be formed, chiefly as a product of burning wood and other biomass. Nitrogen dioxide is hydrolyzed to nitric acid (12). Ammonia can

of animal wastes and plant and animal tissues, application of liquid ammonia fertilizers to farmland, composting and decay of sewage sludge, and industrial processes.

Particulate atmospheric nitrogen results primarily from suspension by wind. Soil disturbance resulting from mining, agriculture, foresting, and construction will contribute significantly to this process.

Most atmospheric deposition is reported in the literature as either total continuous deposition (dust and precipitation) or partial, intermittently collected deposition (precipitation only). Concentrations and unit areal loading rates of various regions are given as representative values in Table 1-2. Variations in the parameters result from variations in both natural background and human activity within the area of meteorological interest.

Surface Runoff. Surface runoff in the urban environment can contain significant levels of nitrogen. Draining of wetlands for land development removes a natural sink for nitrogen (which occurs through the settling of organic solids and denitrification in its sediments). Increased impermeable surfaces ensure rapid conduction of nitrogen to receiving waters, bypassing natural assimilation. Construction and other disturbances create increased quan-

tities of suspended solids (SS) in runoff. These solids generally have a significant particulate organic nitrogen component. Because soil loss from construction can reach over 100 metric tons/ha/yr (45 tons/ac/yr) (19), the portion of area undergoing construction within a watershed need not be great in order to contribute significant nitrogen loading. Urban runoff also generally includes any atmospheric deposition that occurs over its area.

A study of urban sources of nitrogen to Narragansett Bay, Rhode Island, shows typical concentrations from light industrial runoff, roof runoff, and general stormwater runoff to be 0.2–1.1, 0.5–4, and 3–10 mg/L total nitrogen, respectively. Construction site runoff measured in this area yielded 10,000–40,000 mg/L of SS (19). Typical construction site runoff during median flows has been estimated at 1–30 mg/L total nitrogen (20). Other average urban runoff nitrogen concentrations reported from earlier studies are 2.7 mg/L total nitrogen in Cincinnati, Ohio (14); 2.1 mg/L total nitrogen in Washington, DC (21); 2.5 mg/L total nitrogen in Ann Arbor, Michigan (22); and 0.85 mg/L organic nitrogen in Tulsa, Oklahoma (23).

Leakage from failing sanitary sewers, industrial tankage, and septic systems, as well as from illegal hookups and

Table 1-2. Atmospheric Nitrogen: Representative Concentrations and Unit Areal Loadings

Location	Nitrogen Form	Sampling	Measurement	Reference
Representative Concentrations, mg N/L				
Long Island Sound	ammonia	Precipitation	0.13	13
	nitrate and nitrite	Precipitation	0.32	13
Geneva, NY	ammonia plus nitrate-nitrogen	Precipitation	1.1	5
Ottawa, ON	inorganic nitrogen	Snow	0.85	5
	ammonia	Rain	1.8	5
	nitrate	Rain	0.35	5
Cincinnati, OH	total nitrogen	Precipitation	1.27	14
	inorganic nitrogen	Precipitation	0.69	14
Coshocton, OH	total nitrogen	Precipitation	1.17	14
	inorganic nitrogen	Precipitation	0.80	14
Representative Areal Loadings, kg/ha/yr*				
Potomac River	total nitrogen	Precipitation and dust	18.6	15
Lake Huron (northwest)	total nitrogen	Precipitation and dust	11.0	16
Sweden (average)	total nitrogen	Precipitation and dust	10.0	17
Sweden (south)	total nitrogen	Precipitation and dust	15.0–25.0	17
Central Europe	total nitrogen	Precipitation and dust	20.0–30.0	17
Hamilton, ON	total nitrogen	Precipitation and dust	6.5	5
	total nitrogen	Dust	2.6	5
Seattle, WA	nitrate	Dust	0.71	18

* 1 kg/ha/yr = 0.8922 lb/ac/yr

discharges, can collectively account for a significant portion of nitrogen loading in a stormwater collection system (19). Default values used for areal nitrogen loading from low-, medium-, and high-density residential neighborhoods, as given by STORM (the Urban Runoff Model of the U.S. Army Corps of Engineers), are 0.008, 0.031, and 0.028 kg/ha/d (0.007, 0.028, and 0.025 lb/ac/d), respectively. Commercial and industrial area runoff loads are given as 0.237 and 0.234 kg/ha/d (0.211 and 0.209 lb/ac/d), respectively (9). Land uses described as low-to-medium density residential, high-density residential and commercial, and medium- and high-intensity industrial yield average unit nitrogen (25) loadings to storm sewers of 9, 11.2, and 7.8 kg/ha/hr (8, 10, and 7 lb/ac/hr), respectively.

Application of fertilizer to farmlands creates significant nitrogen loadings to ground and surface waters. Controlling factors include application rate of fertilizer, type of fertilizer, irrigation rate, soil drainage, type of plant cover and its nitrogen uptake rate, and degree of tillage. Observed values of nitrogen in runoff and ground water often have a high degree of correlation with these factors (26,27). A survey of the eastern United States characterized stream flow as a function of percent of agricultural land use within each watershed. Mean total nitrogen concentrations for streams with 50, 75, and 90 percent agricultural watershed were 1.08, 1.82, and 5.04 mg/L total nitrogen, respectively (24). A sampling of receiving streams and ground waters from 268 agricultural sites in southeast Nebraska ranged from less than 0.1 to 233 mg/L total nitrogen. At 37 percent of these sites, nitrate-nitrogen levels exceeded 10 mg/L and were often 20–40 mg/L, well above the maximum drinking water limit of 10 mg/L (26). One area in southeast Ireland has five farming districts that have mean artificial nitrogen fertilizer application rates of 47.6–68.2 kg N/ha/yr (42.5–60.9 lb/ac/yr). The mean nitrogen loss to ground water and runoff is 4.1–25.5 kg N/ha/yr (3.6–22.8 lb/ac/yr). The resulting maximum river nitrate concentrations correlate strongly with the corresponding fertilizer application rates. Percent loss values for each farming district also correlate strongly with percent of land area ploughed (27).

Feedlot runoff constitutes a source of nitrogen that has become significant as a result of the increased number of concentrated, centralized feedlots. Ammonium resulting from urea hydrolysis is a major constituent of feedlot waste. Ammonium-nitrogen runoff concentrations may reach 300 mg/L (6,28,29) and organic nitrogen concentrations of up to 600 mg/L (28,29) have been reported. The growing trend is toward feedlot operations as compared to small-farm livestock production. The centralized, more contained nature of feedlots lends itself favorably to collection and treatment, allowing for significant improvements in this area of nitrogen control.

Septic fields are responsible for a significant fraction of the nitrogen load to U.S. ground water. Approximately 25 percent of the population is served by individual home sewage disposal systems (30). Effluent from a typical septic system has a total nitrogen content of 25–60 mg/L. Of this, 20–60 mg/L exists as ammonia and less than 1 mg/L exists as nitrate (9). Another study has characterized a typical septic effluent as containing approximately 7 mg/L organic nitrogen, 25 mg/L ammonium nitrogen, and 0.3 mg/L nitrate-nitrogen (31). A survey of septic fields indicates that rapid nitrification of ammonium-nitrogen takes place under aerobic conditions within the leach field (9). Ammonium-nitrogen is easily exchanged in many soils below a leach field, whereas nitrate remains soluble and is easily lost to ground water. If exchange sites become saturated, as in sandy soils, ammonium breaks through to ground water before it nitrifies. When septic fields dry out in the summer, or are abandoned, much adsorbed ammonium is converted to nitrate, and eventually lost to leaching (31). Natural tertiary POTW treatment systems utilizing soil infiltration or overland flow can typically produce a nitrogen loading to ground water in a manner similar to septic fields. A typical final effluent of this type has a total nitrogen of 3–10 mg/L (9).

The fractional contribution from each of the categories discussed above varies primarily according to the geographical location of the receiving water body, the type and intensity of the development within the region influencing the study site, the population density, and the type of original natural habitat. A number of major studies examine in part the various sources of nitrogen loading to water bodies in the United States. Results of studies of the Long Island Sound, Chesapeake Bay, and coastal waters in Sweden are summarized in Table 1-3 (13,32,33).

1.5 Effects of Nitrogen in Discharges from Wastewater Treatment Plants

Excessive accumulation of various forms of nitrogen in surface and ground waters can lead to adverse ecological and human health effects. This section gives an overview of several effects attributable to nitrogen that can originate from municipal wastewater discharges. One of the major effects has been the direct and indirect depletion of DO in receiving waters. In-stream nitrification directly consumes oxygen, while biostimulation of aquatic plant growth lowers oxygen indirectly when the plant growth dies and undergoes bacterial decomposition. Other impacts can be of major importance in particular situations. These include ammonia toxicity to aquatic animal life, adverse public health effects, and a reduction in the suitability of water for reuse.

Table 1-3. Representative Distribution of Sources of Nitrogen to Chesapeake Bay (31), Long Island Sound (13), and Swedish Coastal Areas (32)

Chesapeake Bay	%	Long Island Sound	%	Swedish Coastal Areas	%
Point sources	23	STPs	43.6	Municipal sewage	19.1
Animal wastes	4	Industry	1.8	Atmospheric deposits on sea surface	9.8
Atmospheric ammonium	14	Atmospheric	11.8	Industry	3.8
Atmospheric nitrate	25	Coastal runoff	6.2	Agriculture	26.3
Fertilizers	34	CSO	1.2	Forest and forestry	23.3
Total	100	Tributaries	35.3	Wetlands	8.4
		Total	100	Other land uses	6.3
				Total	100

Note: Totals have been reached through rounding. CSO = combined sewer overflows; STPs = sewage treatment plants.

1.5.1 Biostimulation of Plant and Algal Growth in Surface Waters

A major problem in the field of water pollution is eutrophication, which is defined as excessive plant growth and/or algae "blooms" resulting from over-fertilization of rivers, lakes, and estuaries. Eutrophication can result in a deterioration in the appearance of previously clear waters, odor problems from decomposing plant growth, and a lower DO level, which can adversely affect the respiration of fish, benthic aquatic animals, and attached bottom plant growth.

Primary conditions that are required for plant or algal growth are adequate macronutrients in the form of nitrogen and phosphorus, sufficient carbon dioxide, and light energy; the absence of any one will limit growth. In special cases, trace micronutrients such as cobalt, iron, molybdenum, and manganese may be limiting factors under natural conditions. Carbon dioxide is very seldom a factor in growth-limiting situations. Sunlight becomes growth limiting in deeper waters because of light extinction or in quiescent or stratified water where excessive algal growth at the surface shields light from the lower levels. Since carbon dioxide and sunlight are virtually impossible to control, their manipulation is not considered a realistic means of limiting excessive photosynthesis.

Nitrogen and phosphorous are typically the two key targets for the control of eutrophication problems. After determining which nutrient, if either, is growth limiting, one must determine if and how the amount of the limiting substance entering the receiving water can be controlled. Under some circumstances, removal of both nitrogen and phosphorus may be undertaken to limit algal growth.

Eutrophication is of greatest concern in lakes because nutrients that enter the water body tend to be recycled within the lake and build up over a period of time (34). A river, by contrast, is a flowing system in which nutrients

are always entering or leaving at any given section. Accumulations tend to occur only in sediment or in slack water, and the effects of these accumulations are normally moderated by the periodic flushing action of floods.

In estuaries and oceans, nitrogen compounds are often present in very low concentrations and may limit the total biomass and the types of species present (34). Thus, upwelling, which brings nutrient-rich waters to the surface, may result in periodic blooms of algae or other aquatic plant life. In some estuaries, discharges from wastewater treatment plants may increase nitrogen concentrations to the level where blooms occur. However, the high dilutions provided by a direct ocean discharge probably eliminates the danger of algae blooms caused by such discharges.

Biostimulation caused by excessive nitrogen loading is considered to be the single largest cause of hypoxia in the Long Island Sound (35). In 1990 the base loading of nitrogen from sewage treatment plants into the Sound was estimated to be approximately 63,600 kg (140,000 lb/d), which accounted for approximately 43.6 percent of the total nitrogen load contributed from all sources. If maximum biological nutrient removal were accomplished in all sewage treatment plants, the resulting maximum effluent nitrogen concentration of 4 mg/L (and effluent BOD_5 (5-day biochemical oxygen demand) of no more than 10 mg/L) would result in a new sewage treatment plant nitrogen loading of 17,770 kg (39,130 lb/d). This would increase the average minimum DO concentration from 1.8 mg/L to approximately 3.3 mg/L in the critical Western Narrows region of the Sound. A major nitrogen removal initiative is being undertaken to upgrade a majority of the sewage treatment plants discharging to the Sound.

While nitrogen in wastewater treatment plant effluents can sometimes contribute to undesirable aquatic growths,

determination of the limiting constituent should be made before the decision is made to require nitrogen removal in the municipal treatment works. It is important to understand that eutrophication is typically a basin-wide issue, particularly in confined water systems. The sources of nitrogen can be widespread, including atmospheric deposition and surface runoff in addition to specific point sources. Thus, it is critical to understand a wastewater treatment plant's contribution to the overall nitrogen load and its significance before considering the imposition of specific nitrogen controls.

1.5.2 Depletion of DO Caused by Nitrification

Just as bacterial decomposition of the carbonaceous organic components of wastewaters depletes DO in the receiving waters, in-stream nitrification of ammonium-nitrogen creates an additional oxygen demand. In specific cases where it is shown that the wastewater ammonium-nitrogen content is a significant factor in the depletion of the receiving water's DO, it is best to accomplish nitrification before discharge.

Table 1-4 gives a simplified example of the impact of providing treatment for nitrification. If conventional biological treatment is utilized to provide 90 percent BOD_5 removal, but no ammonium oxidation (secondary treatment), the effluent will still contain over 100 mg/L of total oxygen demand. This high demand may cause significant oxygen depletion in the receiving water if there is insufficient dilution and the environmental conditions are conducive to nitrification (e.g., a sufficient nitrifying population or time to grow one, pH, temperature, buffering). Although having little effect on organic oxygen demand removal efficiency, accomplishing nitrification before discharge will greatly reduce the residual total oxygen demand discharged to the receiving water.

An analysis of in-stream conditions before and after the implementation of point source treatment upgrades provides examples of the impact of nitrification on in-stream

DO levels (37). Table 1-5 excerpts data regarding three receiving waters that were selected because the treatment plants accounted for the major, or only, point source discharges to the river segment and the specific upgrade encompassed wastewater nitrification.

Figure 1-4 presents DO data from a Patuxent River monitoring station (river mile 70.8) near the sag point downstream of the Laurel Parkway Treatment Plant. The upgraded plant was brought on-line in 1974 with secondary treatment and nitrification. These data, collected at flows near the 7 consecutive day, 10-year low flow (7Q10), indicate an average 1978 summer DO concentration of about 7.6 mg/L in comparison to average 1966 and 1967 concentrations of about 3.7 mg/L (37).

1.5.3 Ammonia Toxicity

Ammonia-nitrogen is the nitrogenous compound most responsible for toxicity effects in fish and other aquatic life. Ammonia dissolved in water exists as an equilibrium of molecular ammonia (NH_3) and ionized ammonium (NH_4^+). Toxicity of aqueous ammonia to aquatic organisms is primarily attributable to the un-ionized, molecular free ammonia form, with NH_4^+ species being relatively less toxic.

The most important conditions that control this equilibrium, and hence the percentage of available molecular free ammonia, are pH and temperature. Figure 1-5 presents the fractions of total ammonia available in the un-ionized form as a function of pH and temperature (38). These are presented within the ranges of pH and temperature normally encountered in continental receiving waters.

Ionic strength of a solution also has a noticeable, but less significant, effect on the percent of un-ionized NH_3. The fraction of ammonia in the un-ionized form decreases with increasing ionic strength in hard water and saline water. In most natural freshwater systems the reduction of percent un-ionized ammonia attributable to dissolved solids is negligible. In saline or very hard waters there will be a

Table 1-4. Effect of Ammonium Oxidation on Total Oxygen Demand of Treated Wastewater Discharge (after 36)

Parameter	Raw Wastewater	Secondary Treatment	With Nitrification
Organic matter, mg BOD_5/L	250	25	20
Organic oxygen demand, mg BOD/L	375[a]	37[a]	30[a]
Organic and ammonia nitrogen, mg TKN/L	25	20	1.5
Nitrogenous oxygen demand, mg NOD/L	115[b]	92[b]	7[b]
Total oxygen demand, mg TOD/L	490	129	37
Percent TOD due to nitrogen	23.5	71.3	18.9
Percent organic BOD removed	—	90	92
Percent TOD removed	—	73.7	92.5

[a] Taken as 1.5 times organic matter.
[b] Taken as 4.6 times the TKN level.

14

Table 1-5. Examples of DO Improvement to River Segments Because of Treatment Plant Upgrades to Nitrification (after 37)

Plant: Receiving Water:		Laurel Parkway[a] Patuxent River		Hurricane[b] Hurricane Creek		Springfield[c] Wilsons Creek	
		Before	After	Before	After	Before	After
POTW effluent	$CBOD_5$, (kg/d)/(mg/L)	159/9.5	17.3/0.45	121/50	11.4/2.1	738/9.8	337/1.6
	NH_3-N, (kg/d)/(mg/L)	128/7.7	5/0.14	38/15.5	3.4/0.64	718/9.5	140/0.68
Total of all sources	$CBOD_5$, kg/d	176	35.5	121	11.4	742	337
	NH_3-N, kg/d	135	32.3	37.7	3.4	718	140
Stream	Average DO, mg/L	5.5	7.9	5.6	6.3	6.4	8.2
	Minimum DO, mg/L	3.8	7.6	2.6	4.5	1.4	6.5
	Maximum $CBOD_5$, mg/L	18.0	<1.0	5.6	5.1	26.0	5.0
	Maximum NH_3-N, mg/L	2.2	0.1	5.4	1.3	22.2	<1.0

[a] Before = study period before 1968; After = study period after 1978
[b] Before = study period before 1972; After = study period after 1981
[c] Before = study period before 1968; After = study period after 1979
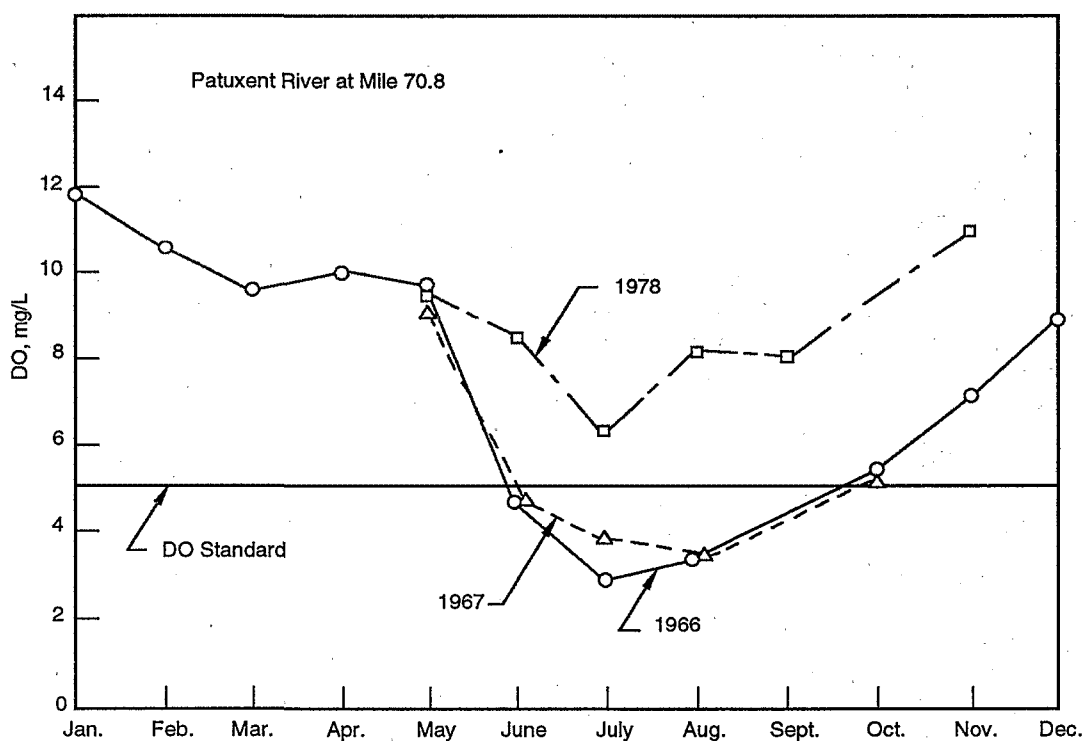
June, July, August & September

	Mean, mg/L		
	1966-67	1978	% Change
DO	3.7	7.6	+ 108%

Figure 1-4. Time history data analysis for main stream of Patuxent River (station: PXT0708 State of Maryland) (from Reference 37).
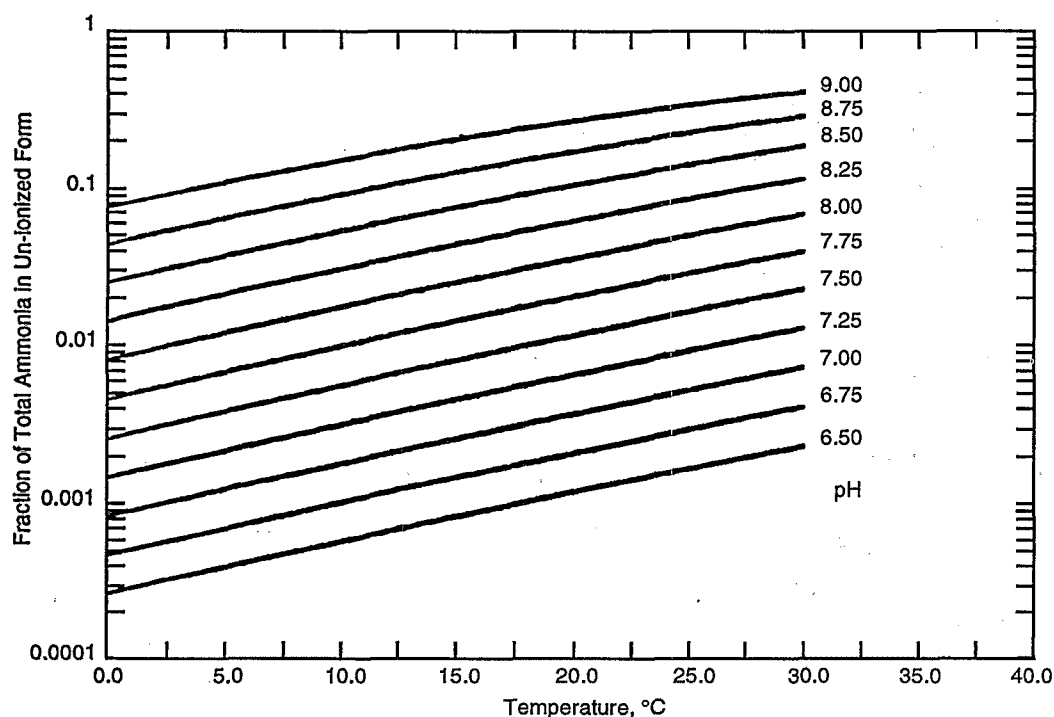
Figure 1-5. Effect of temperature and pH on un-ionized ammonia (from Reference 38).

small but noticeable reduction in un-ionized ammonia fraction (38).

EPA has assembled extensive research to support its ambient water quality criteria for ammonia in freshwaters (38) and saltwaters (39). For both fresh and salt water the majority of research studies show acute toxicity effects for salmonid and nonsalmonid fish species between 0.1 and 10 mg/L un-ionized ammonia. The EPA criteria for ambient water quality, as well as modified-state criteria, give both maximum total and un-ionized ammonia levels as a function of pH and temperature. The maximal one-hour average in-stream concentrations of un-ionized ammonia permissible in a three-year period are all under 1 mg/L. The maximal four-day average concentrations for the same are all under 0.1 mg/L. Figure 1-6 is a graphical display of EPA chronic toxicity criteria for ammonia, with salmonids absent. Supporting research is reported also for aquatic invertebrates and plants (38).

Free un-ionized ammonia toxicity effects as a function of temperature and pH are observed in eutrophic lakes. Lake Onondaga, in New York, is a representative example (40). With warm-weather stratification of lake waters, the DO content of the bottom waters, or hypolimnion, decreases with bacterial respiration and the lack of circulation with aerated surface waters. Total ammonia, from the decomposition of decaying organic sediment, accumulates as nitrification decreases under growing anoxic conditions. A warm-weather vertical profile of a typical lake shows high total ammonia/low nitrate concentrations

in the hypolimnion and low total ammonia/high nitrate concentrations near the surface.

Compared with other systems, Lake Onondaga maintained relatively high total ammonia concentrations in the epilimnion, or surface waters, throughout the spring-to-fall study period in 1988. The lake experienced large fluctuations in algal biomass, reflecting strong variations in the net algal growth rate. Periods of oxygen supersaturation and elevated solution pH occurred as a result of photosynthetic oxygen production and carbon dioxide consumption. Because of these elevated pH (and temperature) levels at the surface, the distributions of free un-ionized ammonia contrasted strongly with those for total ammonia. Maximum free un-ionized ammonia concentrations were noted in the surface waters, despite the fact that the highest total ammonia concentrations occurred in the bottom waters. A vertical profile of the observed free un-ionized ammonia with depth is shown on Figure 1-7, excerpted from the Lake Onondaga study. The figure also presents the computed final chronic values (FCV) and one half the final acute values (0.5 FAV) from the EPA Water Quality Criteria as a function of pH and temperature (note that 0.5 FAV is shown because the criterion states that average one-hour un-ionized ammonia concentrations must be less than one-half the FAV). The elevated free un-ionized ammonia levels resulted in continuous contravention of the chronic ammonia toxicity criteria for nonsalmonid fish and less frequent contravention of the acute toxicity criterion for nonsalmonids.
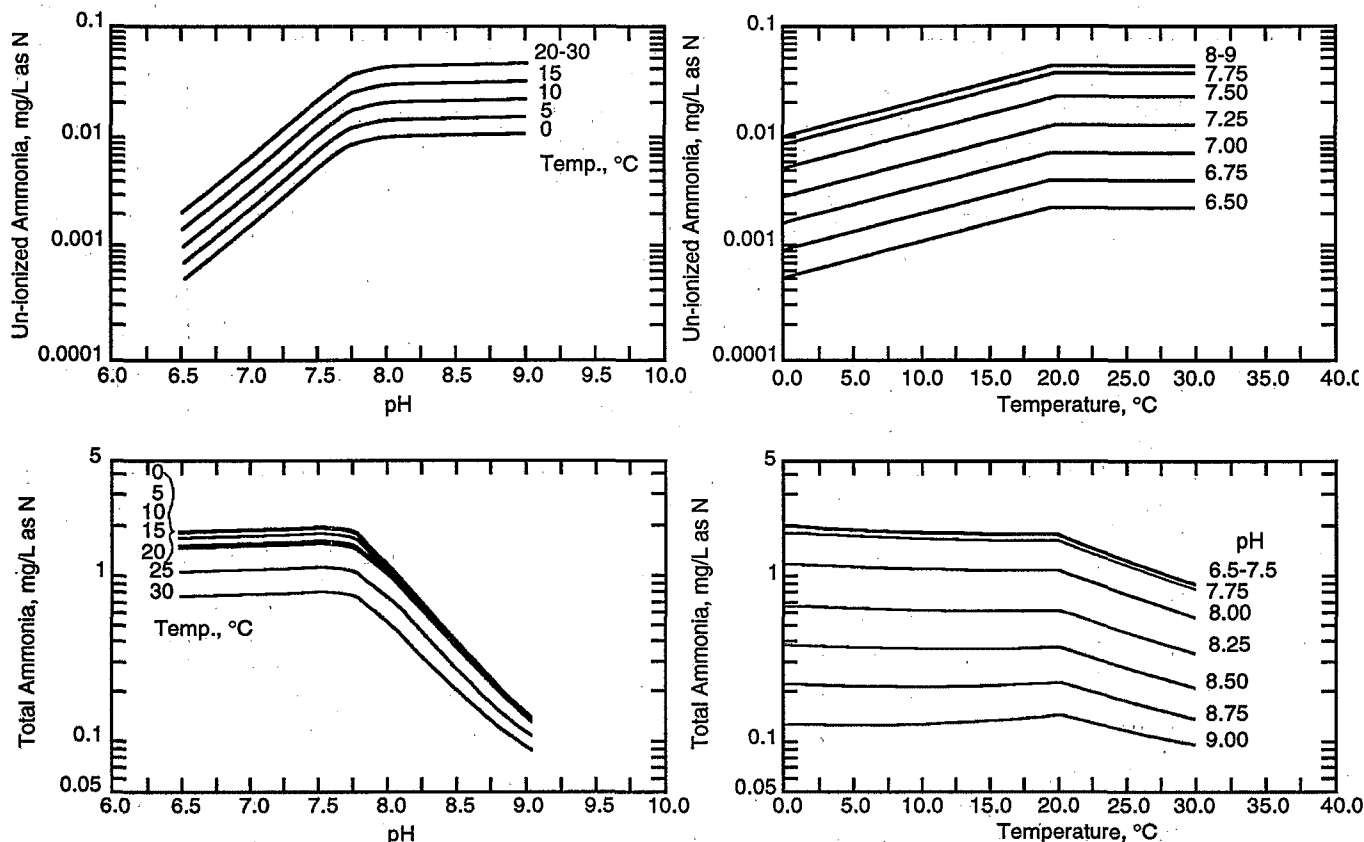
Figure 1-6. EPA chronic criteria for ammonia (salmonids absent) (from Reference 38).

1.5.4 Public Health

Nitrate and nitrite nitrogen constitute a public health concern, related primarily to methemoglobinemia (infantile cyanosis) and carcinogenesis. Methemoglobinemia is a disease primarily affecting infants and is often described by the lay term *blue babies*. The acute toxicity of nitrate occurs as a result of its reduction to nitrite, a process that can occur under specific conditions in the stomach and saliva. The nitrite ion formed oxidizes iron in the hemoglobin molecules from the ferrous to the ferric state. The resulting methemoglobin is incapable of exchanging oxygen, and anoxia or death may occur if the condition is left untreated. Suffocation is often accompanied by a bluish tinge to the skin. Most reported water-related cases of infantile methemoglobinemia have been associated with the use of water containing more than 10 mg/L nitrate-N. Thus, drinking water guidelines and standards recommend that the concentration of nitrate in drinking water should not exceed 10 mg/L (28,41,42). This standard is exceeded most often in shallow wells in nonsewered rural areas where septic systems are the dominant method for sewage disposal.

Carcinogenesis (gastric cancer, in particular) has been associated with the ingestion of N-nitroso compounds.

Nitrites (indirectly from nitrates) can react with amines and amides to form nitrosamines and nitrosamides. The epidemiological evidence suggests that high nitrate ingestion may be a contributing factor in gastric cancer. There appears to be little information available to draw any conclusions about any other human cancer and high levels of nitrates in drinking water (41–43).

1.5.5 Reuse of Wastewater

Wastewater reclamation has traditionally been practiced for pollution abatement, although the benefit derived from supplementing available water resources has become increasingly important. The uses of reclaimed water include indirect potable reuse, agricultural reuse, urban landscape irrigation, industrial reuse, ground-water recharge, and potable reuse. The primary obstacle to reclamation is the removal of pathogens. Nitrogen removal becomes more of a concern when reclaimed water is ultimately intended for supplementing municipal drinking water supplies. Although biological processes are generally recommended for nitrogen removal from wastewaters, multistage tertiary treatment often reduces solids and carbonaceous substrate to an extent where physiochemical processing is preferred for nitrogen polishing. The 44-L/s (1 mgd) Demonstration Plant operated by the Metropolitan Denver Sewage Disposal District No. 1 re-

17
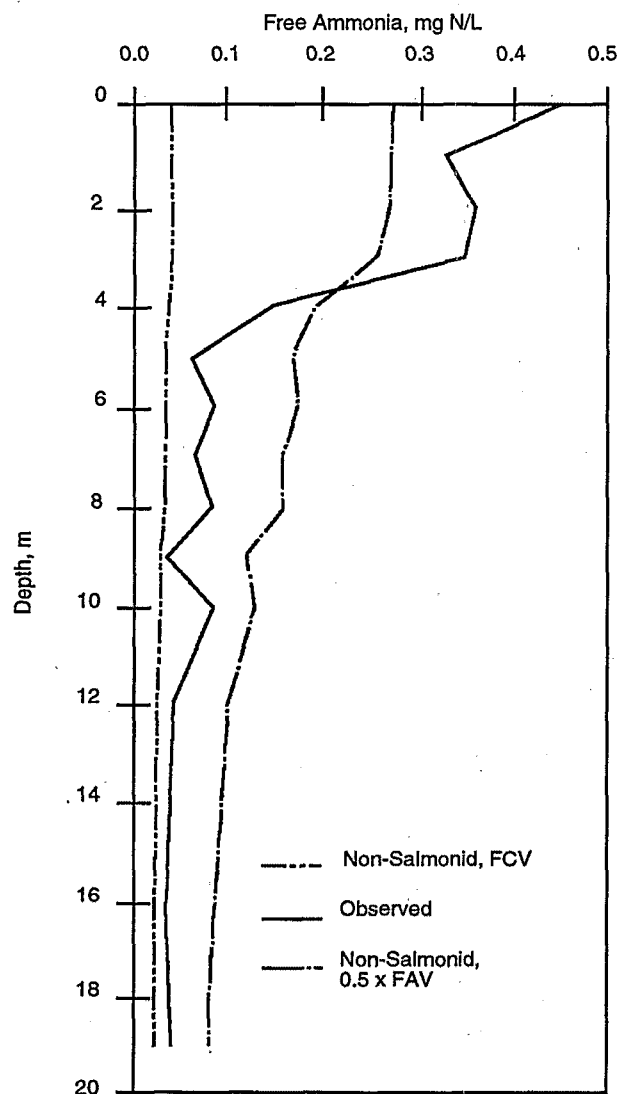
Figure 1-7. Depth profiles of NH₃, FCV, and 0.5 x FAV in Onondaga Lake for June 19, 1988 (adapted from Reference 40).

moves nitrogen in the form of ammonium by passing the water through a naturally occurring zeolite media (clinoptilolite). Sodium chloride is used to regenerate the zeolite media. The ammonia is recovered from the regenerant solution in the form of ammonium sulfate and sold as a commercial-grade fertilizer (44).

Potable reuse is considered indirect when reclaimed effluent is discharged to a surface water supply. The Upper Occoquan Sewage Authority (UOSA) removes ammonium by ion exchange. The ion exchange regenerant stream undergoes regeneration in a closed-loop ammonia stripping and adsorption process that yields a 40 percent ammonium sulfate solution for resale as an agricultural fertilizer (44). Ammonia removal for the purpose of ground-water recharge is accomplished at the Orange County Sanitation District reclamation plant known as Water Factory 21. Tertiary reclamation proc-

esses reduce the secondary effluent from 45 mg NH_4^+-N/L to 4.7 mg/L. After blending with lower concentration sources, the final product contains an average of 0.9 mg NH_4^+-N/L to satisfy an injection requirement of 1.0 mg/L (45).

Nitrogen limits on water for reuse in agricultural and urban landscape irrigation are not a factor, providing that there are no secondary concerns for augmenting the nitrogen load to ground water. On the contrary, there may be a benefit in having residual ammonium and nitrate in water used for irrigation. Nitrogen requirements for industrial reuse vary widely according to the particular process for which it is used. A considerable amount of water in this category is used for cooling purposes, in which case the concern is primarily total dissolved solids, of which nitrates are typically a small component.

1.6 Establishing Nitrogen Limits for Wastewater Discharges

The Federal Water Pollution Control Act of 1965 began a series of environmental legislative reforms that initiated a consistent approach to pollution control based on water quality and beneficial use goals. National technology-based standards were established, moving all wastewater treatment facilities to secondary levels, at minimum. Requirements for advanced treatment, such as ammonia removal or total nitrogen removal, reflected specific water quality needs for the given receiving water. National water quality goals have influenced the development of advanced treatment technologies, especially in the area of nutrient control.

Various beneficial uses of a water body include potable water supply, support and propagation of fish and wildlife, recreational activities, agricultural irrigation, industrial supply, navigation, and shipping. Excess nitrogen in its various forms will have adverse effects on a receiving water (as discussed earlier in Section 1.5), anyone of which can impact the desired use of the system. Where appropriate, limits are imposed on the discharge of nitrogen (in addition to conventional parameters such as BOD and SS) from point sources such as wastewater treatment plants as a means of controlling the water quality and ensuring the long-term beneficial use of a receiving water. These "wasteload allocations" are generally based on water quality criteria or standards established for the specific receiving water.

The 1987 re-enactment of the Federal Clean Water Act allows the states to designate water use and to establish water quality standards (WQS). These generally follow or are more restrictive than the guidelines promulgated by EPA; although they are defined based on protocols set by EPA. Through research both in and outside of EPA, data are compiled from which recommendations are made regarding the biological and chemical conditions necessary to sustain or achieve a water body's designated use. In the cases of ammonia toxicity to aquatic biota and nitrate effects on human health, the mechanisms are well quantified and understood and generally nonspecific with regard to receiving waters. As such, criteria are established and broadly implemented (these are discussed in Sections 1.5.3 and 1.5.4). The impacts relating to in-stream nitrification and biostimulation tend to be more site-specific. Determination of their causes and the implementation of controls often require water quality studies and modeling evaluations specific to the water body.

Technology-based standards do not exist for nitrogen, as they do for SS and BOD, because of the varied nature of nitrogen's effect on receiving waters. As such, state agencies, with guidance from EPA, will generally use water quality modeling approaches to determine appropriate point source wasteload allocations that will sustain ambient water quality standards. If nitrogen control (nitrification or nitrogen removal) is needed within a water system, then a quantitative analysis of all point and nonpoint sources is required to determine the appropriate treatment needs for the municipal discharge.

To ensure that current discharge limits and waste load allocations applied to each water body effectively maintain the water quality standards set for that body, states periodically assess their condition. This is generally done every three years, as mandated by the Clean Water Act. States usually prioritize their waters for evaluation and assess existing data for each. If data are lacking and the waterway is a priority, a Water Body Survey and Assessment is conducted. If it is found that existing discharge limits are not effective in maintaining the beneficial use of the water, improvements in treatment and controls are generally required. The state may also propose a downgrading of the designated use of the receiving water if it finds that natural conditions and economic constraints make maintaining the beneficial use impractical. This, however, is less common and protocols for changing to a less-protective use require a rigorous evaluation that includes public participation. The process of water quality standards review and revision is outlined in Figure 1-8.

In summary, state-set ambient water quality standards for nitrogen are translated via wasteload allocation modeling methods into water quality-based permits for POTWs. These permits serve as the basis for water quality-based construction funding decisions. Figure 1-9 outlines the major elements of the water quality-based standards-to-permits process.
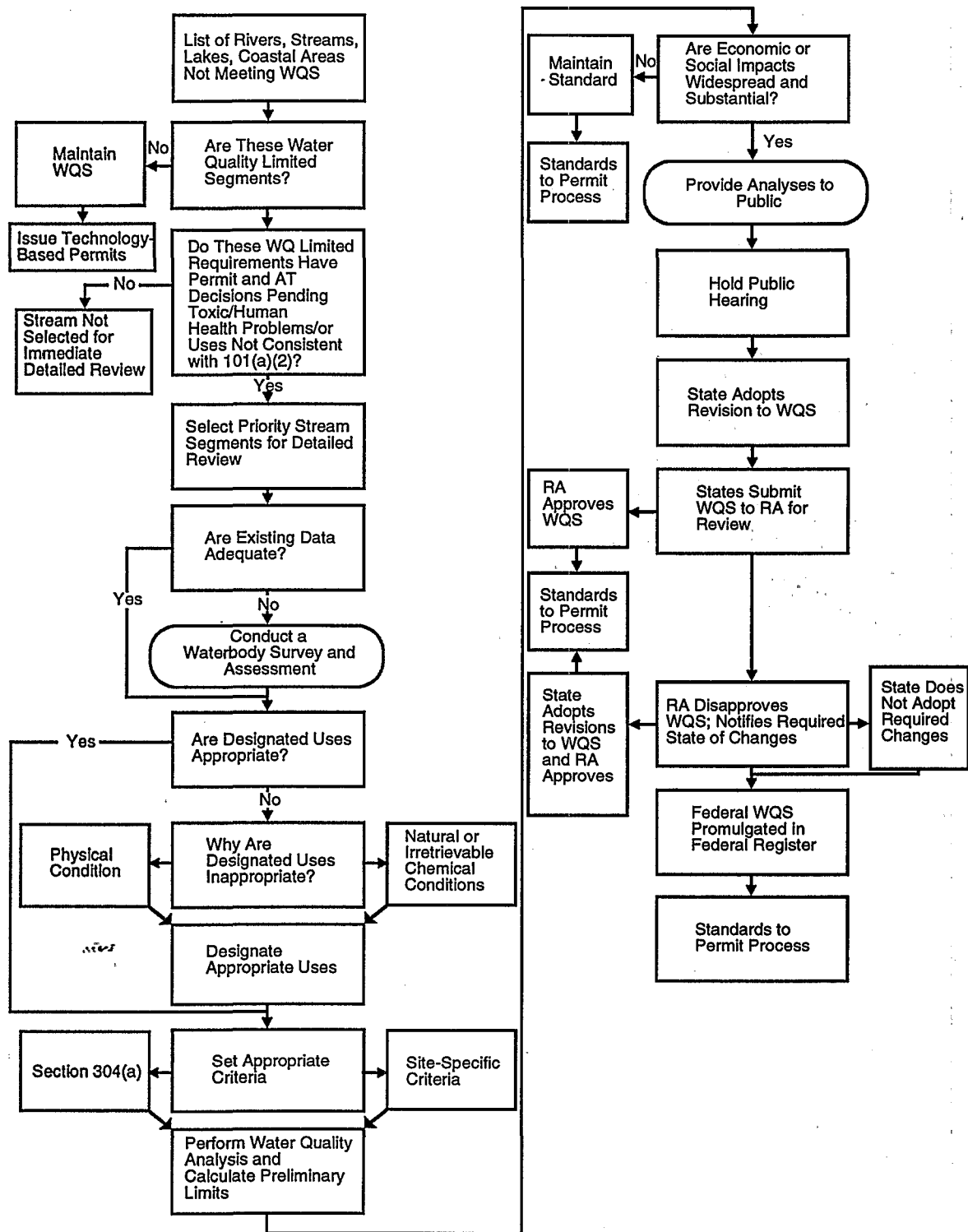
Figure 1-8.   EPA water quality standards: Review and revision process (from Reference 46).
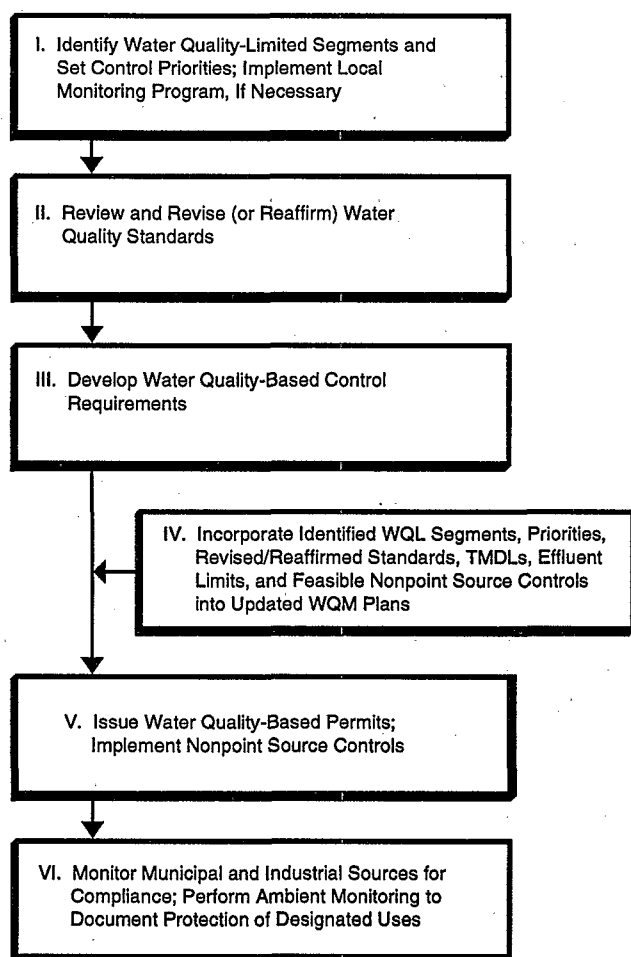
Figure 1-9. Major elements of the water quality-based standards to permits process (from Reference 46).

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When an NTIS number is cited in a reference, that document is available from:

National Technical Information Service
5285 Port Royal Road
Springfield, VA 22161
703-487-4650

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Chapter 2
Total System Design with Nitrogen Control

2.1 Introduction

For many readers, this is one of the more important chapters in this manual. It is directed to all decision-makers involved in the assessment, selection, and design of a nitrogen control strategy at a municipal wastewater treatment facility.

The overall intent of this chapter is to convey the message that there is no universal response to a nitrogen control need. The right unit process selection, rather than standing alone, is influenced by all things that precede and follow in the overall scheme of the treatment works. Process selection must respond to the facility's wastewater and residuals management objectives, under the constraints imposed by both the natural and social environments.

This chapter:

- provides summary oversight and detail at a level and in a form that are convenient for frequent reference,

- introduces and compare the various nitrogen control technologies and fundamental aspects of their performance, and

- describes sound practices and linked processing considerations that will assist the user in the application of nitrogen control technology and with related problem solving within the context of the integrated treatment works.

The information presented in this chapter focuses on avoiding the fundamental mistakes often encountered in the first 5 to 10 percent of the design, which, if continued through the remainder of the project, can result in a facility that either fails to meet its design intent or is grossly oversized.

The interested reader will be best served by reading this chapter for overall understanding before referring to the more specific and detailed nitrogen control technology material contained in the balance of the manual. After reading the more detailed design chapters, rereading this chapter may be appropriate before entering into detailed process design. After design completion, the material contained in Section 2.2 can be used to check some of the design decisions.

Those readers desiring more information than found in this manual are referred to the recently revised Water Environment Federation's Manual of Practice (MOP) No. 8, *Design of Municipal Wastewater Treatment Plants* (1). The three introductory chapters of that publication provide greater understanding of design approaches and issues, wastewater characteristics, and the decision-making that precedes detailed design.

2.2 Summary Checks for Process Selection and Design

This section summarizes a variety of material and will serve as a convenient reference. The section begins by focusing on the most frequently encountered problems at small wastewater treatment plants. Then, with an emphasis on nitrogen control aspects of the subject, the reader is provided with a brief narrative and tabular summary of general guidance checks for wastewater characterizations and waste solids management schemes, the stoichiometry of the frequently encountered reactions of interest, and an oversight review of the proposed process design.

2.2.1 Overview

Successful wastewater treatment is dependent on operator understanding, responsible administration, and sound design. Failure of any agent of successful wastewater treatment (i.e., operator, administrator, or designer) to respond adequately to his charge inevitably results in process upset and eventual failure.

The designer can mitigate the failure opportunities at all treatment facilities by:

- selecting tolerant, flexible treatment processes with conservatively designed, responsive sludge processing and disposal schemes,

- urging and providing for operator training, and

- working with the administrator to ensure adequate support for the needs of the facility.

Small wastewater treatment plants are more likely than larger plants to be underfunded, understaffed, and not always reached by the professional networks that can offer troubleshooting advice. Table 2-1, taken from the recently revised MOP 8 (1), summarizes the findings of a recent EPA survey of 150 small plants (≤44 L/s [1 mgd]) with debilitating problems (2).

Table 2-1 reveals that the three agents responsible for successful wastewater treatment plant operation also share blame. All would do well to remember the seven major conclusions and recommendations derived from the EPA study.

2.2.2 Wastewater Characterization and Waste Solids Management

The three biggest errors in design of nitrogen control municipal wastewater treatment facilities are the same as those encountered in any treatment application:

- Improper characterization of the influent wastewater,

- an inadequate and/or improper solids processing and disposal scheme, and

- failure to consider transient sludge processing recycles in relation to the buffering capacity of the wastewater and the processes found in the liquid processing train.

Sound design uses a feed forward (wastewater characteristics), feed backward (solids disposal and effluent limitations) approach to develop the liquid and solids processing trains that are compatible with the anticipated staffing plan and the constraints or opportunities that are established in the natural and institutional environments. The proper design provides the least expensive, implementable solution that results in balanced media (i.e., air, land, and water) impacts and that complies with the current and anticipated institutional needs of the public, the owner, and the regulatory bodies.

Common sense and detailed process understanding go hand in hand in soundly based process applications. Table 2-2 lists some of the wastewater characterization and solids management checks that may be employed by both the process designer and the reviewer to ensure a successful application of nitrogen control technology in municipal applications.

Two-thirds of Table 2-2 is devoted to influent wastewater characterizations. Characterization errors may be inadvertent (e.g., didn't know any better or trusted an invalid data base) or conscious (e.g., overquantification of the influent flow and pollutant load to safeguard against failure).

The effluent standards of the plant (pollutant limit and time interval) dictate the needed influent characterizations. These characterizations then feed forward into the design of those processes needed to meet the effluent standards. If the plant has a maximum month and week per year effluent pollutant restriction (corresponding to a de-

sign reliability of 92 and 98 percent, respectively), influent wastewater pollutants should also be characterized at these conditions to permit the establishment of the controlling condition for facility design.

However, even the best liquid processing design is meaningless if the solids processing train is inappropriate, improperly sized, or unreliable. The balance of Table 2-2 is devoted to the issues that are found therein and the recycles that are generated by the solids processing train.

Solids management decisions, beginning with the ultimate disposal concept, feed backward into the facility design and, practically, can have much to do with the sound selection of the liquid processing technology. This is especially the case when nitrogen control is a processing objective. Failure by the designer and owner to appraise realistically and provide for adequate ultimate solids disposal places the entire treatment complex at risk.

The recycles returned to the liquid processing train, and the waste solids from it, integrate the treatment works so that "everything is connected to everything else." Too high a return of SS from the solids processing train can cause a plant's solids residence time (sometimes referred to as the mean cell residence time), as determined by dividing the biomass in the reactor by the mass of SS removed (wasted and lost in the effluent) per day, to drop below the acceptable value for retention of the nitrifying organisms. An additional result is the spin (continuous recirculation) of previously generated solids, which taxes the processing capability of all processes up to and including the one that was the source of the recycles. The discontinuous return of a soluble nitrogenous recycle imposes a special load that must be anticipated in design and buffered by the processes and alkalinity of the wastewater to avoid undesirable nitrogen excursions in the plant's effluent.

2.2.3 Stoichiometry

Wastewater treatment plant design in general and nitrogen control concepts specifically involve a variety of specific stoichiometric reactions and more uncertain process assumptions. Table 2-3 summarizes the stoichiometric reactions that are frequently employed and encountered in the design of nitrogen control systems. Additional information regarding these approaches, and all of the stoichiometric reactions, are provided in the referenced sections of the manual. The stoichiometric constants are described to two significant figures; greater precision is unnecessary.

Table 2-3 shows that the nitrogen control reactions can deplete and add alkalinity. Alkalinity control is important in the process design to avoid low pH attenuation of the nitrification reaction. The alkalinity provided from a variety

Table 2-1. Survey of 150 Small Plants with Debilitating Problems (Adapted from References 1 and 2)

MOST FREQUENTLY OCCURRING PERFORMANCE LIMITING FACTORS (PLFs)
(From Survey Summary of Five Most Frequently Occurring PLFs by 10 Regions)

| | | Likely Principal Causative Agent | | |
No. of Regions	Performance Limiting Factors	Operator	Administrator	Designer
10	Poor Operator Understanding/Application of Process Control	X		
9	Solids Handling and Sludge Disposal		Uncertain	
8	Infiltration/Inflow		Uncertain	
6	Staffing (e.g., too few staff, low pay, turnover)		X	
6	Laboratory Capability for Process/NPDES Testing		Uncertain	
4	Process Design Errors (e.g., clarifiers, aerators, disinfection)			X
4	Municipal Support (administrative and technical)		X	
3	Preventive Maintenance Program		Uncertain	

SURVEY CONCLUSIONS AND RECOMMENDATIONS

- Conclusion No. 1 -*Activated sludge may not be a good design choice for many small plants.*

  – Give fuller consideration to simpler, more tolerant treatment processes (e.g., fixed media and natural systems) that are less dependent on highly skilled operators.

  – Select a treatment technology based on realistic appraisal of all costs (including conservative estimates of sludge quantity and concentration for sludge treatment and disposal, staff salary, recruitment and training, equipment maintenance and replacement, and administrative costs).

- Conclusion No. 2 -*Plant inflexibility undermines operability.*

  – Designers should conscientiously build flexibility into systems (e.g., piping configurations, redundant unit processes, variable speed pumps for wasting and return, aeration equipment, and equalization tanks for I/I extremes and/or bleed-back of discontinuous sludge processing recycles to the liquid processing train).

- Conclusion No. 3 -*Small plants have front- and back-end problems with process design.*

  – Pumps, piping, and aeration systems should be designed to accommodate increased solids and rags in the system when primary treatment is not provided.

  – Operators should be made aware of the need to remove floating debris that passes primary screening.

  – Designers should consider finer bar screens, especially when primary sedimentation is not provided and, once screenings and floatable material are removed from the liquid processing train, provide practical facilities to facilitate their permanent removal instead of their internal recycle (and buildup).

  – Solids handling facilities should have the capability of properly stabilized liquid sludge removal and disposal.

- Conclusion No. 4 -*Heavy loads can confound both skilled and unskilled operators.*

  – Community administrators and design engineers should frankly discuss and agree on realistic loadings for the facility in the planning process (a conservative, if not skeptical, design approach should be taken to accommodate I/I and industrial loadings).

- Conclusion No. 5 -*Staffing difficulties aggravate poor performance.*

  – Administrators should seek to attract and maintain a better staff through increased operator status and visibility using at least one (or nearly so) full-time position with a salary comparable to other critical municipal functions (e.g., the police chief) and with reasonable authority for budgeting, purchasing, hiring and firing. Administrators should also provide reasonable opportunities for training and certification.

- Conclusion No. 6 -*Plant budgets and user charges may be too low.*

  – Better fiscal management must start with a separate budget for the treatment plant that includes a sinking fund to cover replacement of major equipment, and that supports adequate staff salaries as well as training and required certification courses.

- Conclusion No. 7 -*Municipal support is a subtle but vital need.*

  – Outreach and information transfer must be applied to increase community support; consider making the treatment plant into a multi-use facility that accommodates recreational facilities and shares offices and building space with other community agencies and organizations.

Table 2-2. Wastewater Characterization and Solids Management Checks for the Design of Municipal Wastewater Treatment Facilities with Nitrogen Control Technology

RAW INFLUENT WASTEWATER CHARACTERISTICS

- Per Capita Pollutant Generation Rates

  - BOD_5 and SS = 0.08–0.11 kg/capita/d (0.18–0.25 lb/capita/d)

    - Low values are favored with strictly domestic service and absence of household garbage grinders.

    - Quantification lower than cited value should be confirmed (may be encountered with lower income communities), and may be symptomatic of upsystem diversions and/or poor sampling or analysis.

    - Quantification higher than cited value should be confirmed and may be symptomatic of unquantified plant recycles, unregulated septage dumping, sampling programs that exclude weekends, poor sampling or analysis, incorrect flow identification (e.g., using influent and recycle flows with raw wastewater characteristics), industrial or commercial discharges, and significant daily transient population. If confirmed as industrial or institutional, determine seasonal, weekly, and daily waste discharge characteristics and plan for the future, as appropriate. Watch out for sampling schedules at small plants that may be limited to only one to five days per week with sampling only during staffed hours. Manual composite samples may not be true composites. Remember that industrial and commercial releases are typically less on weekends and holidays (i.e., after the end of the Friday shift cleanup).

    - Perform long-term inert SS mass balance check around whole plant (influent against effluent SS and net waste solids). See also waste solids production.

    - Sustained low-flow period may give more reasonable characterization than high-flow period given sewer system overflows and losses. However, high-flow period may experience load from previously deposited, now resuspended solids. See also maximum and minimum characterizations.

- Pollutant Relationships

  - BOD_5/COD = 0.45–0.55

    Lower values may indicate a fair degree of stabilization occurring in the sewers (enhanced by steep slopes, aerobic conditions, higher temperatures), or high levels of I/I (influx of more refractory organics); or attributable to nonacclimated seed or poorly biodegradable industrial waste. Higher values may indicate fermentation in long residence time anaerobic sewers, false BOD_5 positives due to sulfide, presence of nitrifiers due to recycles, or high levels of soluble biodegradable industrial waste.

  - VSS/SS = 0.7–0.8

    Higher values are favored with domestic wastes. Lower values are often encountered in combined sewer areas. VSS-to-SS ratios less than 0.7 should be confirmed and are indicative of the routine receipt of partially stabilized wastes (septage), water plant sludges, industrial wastes or pronounced precipitation induced inflow.

  - SS/BOD_5 = 0.8–1.2

    See preceding paragraph.

  - Soluble BOD_5/Total BOD_5 = 0.35–0.45

    See preceding paragraph. Higher values may be indicative of industrial waste. High values dictate special concern with filamentous bulking because of high immediate stabilization and DO stresses in suspended growth system.

  - COD/VSS = 1.3–1.8

  - Particulate BOD_5/Particulate COD ≤ Total BOD_5/Total COD

    Soluble phase is typically the most biodegradable.

  - TKN/BOD_5 = 0.1–0.2

    Higher values may be indicative of industrial waste or the presence of ammonium from solids processing recycles introduced before sampling the influent. Lower values may be indicative of nutrient-deficient industrial waste.

- Maximum And Minimum Characterizations

  - *Flows*

    Guard against the inadvertent inclusion of plant recycles quantified with the raw sewage determination because of location of the flow meter. Maximum flows may be constrained by sewer system or headworks diversions before the flow meter. Snow melt can give high flows. Minimum flows may be limited by seasonal infiltration.

  - *Pollutants*

    Guard against poor sampling and/or analysis; this becomes progressively more important as maximums and minimums are identified. Look for similar trends of wastewater constituent ratios as a form of data validation. Validate observations by determining if operating solids levels and waste solids production values follow reported pollutant peaks. Delete any

erroneous result from average determination. Analyze several years of data to confirm. Common errors are failure to completely filter SS samples and overflowing sampling bottles, which yield high SS. Samples with high settleable SS may yield low BOD_5 values because settled solids are not suspended throughout the duration of the test. Believe COD values more than BOD_5 determinations. Make sure that calculated masses are determined with concentration and flow of event.

— *Alkalinity*

Strongly influenced by native water and water treatment practices, as well as by magnitude of inflow and infiltration. May be influenced by sea water intrusion in low-lying coastal areas. Note: Inexperienced designers often ignore alkalinity when designing nitrogen control facilities, especially when planning on the discontinuous return of unoxidized soluble nitrogen-laden recycles from the solids processing train.

WASTE SOLIDS PRODUCTION, PROCESSING, AND DISPOSAL

● Net Waste and Effluent SS

For most municipal wastewaters with and without primary treatment but without enhanced biological or chemical phosphorus removal, where applied $(BOD_5 + SS) \div 2 = A$, the combined net waste and effluent SS are typically on the order of the following "rule of thumb" estimates below. (Additive inert solids for enhanced biological or chemical phosphorus removal is 0.02 to 0.04 times ΔP, where ΔP is the enhanced P removal in mg/L.)

	Multiplier of A
Carbonaceous Stabilization (Oxidation)	0.9–1.1
with aerobic or anaerobic digestion of total sludge mass	0.7–0.9
Carbonaceous and Nitrogenous Oxidation	0.8–1.0
with aerobic or anaerobic digestion of total sludge mass	0.7–0.9

● Ultimate Disposal Scheme

If disposal outlets are not in the owner's total control, or processed sludge storage is inadequate for at least several months of storage prior to ultimate disposal, the residuals management scheme is at risk and the liquid processing scheme may be subject to excessive solids-laden reycles; this may upset the performance of one or more unit processes of the liquid processing.train.

● Solids Processing

— *Nitrogen Recycles*

Soluble nitrogen causes the most severe nitrogen impact. All solids, whether in process or in storage, undergo hydrolysis (liquification) and contribute soluble nitrogen. Nitrogen solubilization is most severe in thermal sludge conditioning and digestion (aerobic, anaerobic, and composting); neither contribute a favorable biodegradable carbon-to-nitrogen ratio such that the recycled nitrogen is completely removed by biomass synthesis after the flow is returned to the aeration basin. Aerobic digestion can also contribute ammonium-nitrogen, when nitrogen oxidation in the aerobic digester is limited by the low pH (typically from 5 to 6) resulting from an unfavorable soluble nitrogen-to-alkalinity ratio. Discontinuous solids processing subsequent to the process that causes solubilization exacerbates the stress caused in the liquid processing train by the soluble nitrogen. The more discontinuous the solids processing/dewatering process, the greater the buffer or treatment reserve needed in the liquid process train. Note: Soluble refractory organic nitrogen, such as encountered with discontinuous sludge processing or septage, may seriously affect the ability of a plant to achieve stringent total nitrogen effluent requirements.

— *SS Recycles*

Sedimentation tanks in the liquid processing train are best used for clarification, not thickening. Most optimum solids processing is achieved by attempting to concentrate the waste solids in processes that are least sensitive to flow (e.g., gravity thickening and dissolved air flotation) before using other more volume-dependent processes (e.g., digestion, centrifugation, and belt thickening and dewatering). This avoids SS washout and excessive SS recycle as well as inadequate mainstream capacity resulting from the failure to achieve the anticipated solids concentration in the liquid processing train. Care should be exercised in assuming that gravity thickeners have too much storage (i.e., undersizing) as SS captures can change from greater than 95 percent to disastrously low values in a matter of hours.

Table 2-3. Stoichiometry of Nitrogen Control and Other, Often Related Reactions

	MANUAL SECTION

CONSIDERATION: NITROGEN CONTROL REACTIONS

Biochemical Nitrification   3.2.1

- 4.6 mg oxygen required/mg nitrogen oxidized
- 7.1 mg $CaCO_3$ alkalinity depleted/mg nitrogen oxidized
- 0.10–0.15 mg net volatile solids/mg nitrogen oxidized

Biochemical Denitrification   4.2.2.1, 4.2.2.2

- 2.9 mg oxygen liberated/mg nitrate nitrogen reduced; there are 1.5 mg COD/mg methanol (CH_3OH)
  Note: Sufficient substrate (COD) must be added to satisfy nitrogen reduction and synthesis needs, typically about 1.5 times stoichiometric predictions.
- 3.6 mg $CaCO_3$ alkalinity is recovered/mg nitrate nitrogen reduced
- Same to slightly lower net volatile solids/COD removed as with any biological system (methanol is preferentially oxidized as opposed to synthesized); yields can be lower if an aerobic stabilization step is provided
  Note: COD removed is the total amount of COD oxidized and synthesized.

Breakpoint Chlorination   2.5.2.3

- 7.6 mg chlorine/mg ammonium nitrogen
  Note: Practice typically requires values 1.1–1.3 times stoichiometric predictions when treating biologically stabilized wastewaters. Values may be appreciably greater when processing raw or partially stabilized wastewaters because of competing side reactions with the SS, organics, and nitrite nitrogen.
- 1.4 mg $CaCO_3$ alkalinity loss/mg chlorine added (1 mg dissolved solids added/mg chlorine added)
- NaOCl is buffered (with caustic) and will not cause an alkalinity depletion (1.7 mg dissolved solids added/1 mg chlorine [in NaOCl] added)

CONSIDERATION: OTHER, OFTEN RELATED REACTIONS

Chlorine Demand Due to Incomplete Nitrification or Denitrification   2.8.5

- 5.1 mg chlorine/mg nitrite nitrogen, yielding nitrate nitrogen

Dechlorination

- 0.9 mg SO_2/mg chlorine (expressed as Cl_2)
  Note: Practice requires values 1.1–1.2 times stoichiometric predictions. (2.4 mg $CaCO_3$ alkalinity depleted/mg SO_2 added)
- 1.5 mg $NaHSO_3$/mg chlorine (expressed as Cl_2)
  Note: Practice requires values 1.1–1.2 times stoichiometric predictions (1.4 mg $CaCO_3$ alkalinity depleted/mg $NaHSO_3$ added)

Alkalinity Sources

- 1.8 mg $CaCO_3$ alkalinity added/mg CaO (quicklime) added
- 1.4 mg $CaCO_3$ alkalinity added/mg $Ca(OH)_2$ (slaked lime) added
- 1.2 mg $CaCO_3$ alkalinity added/mg NaOH (caustic) added
- 0.9 mg $CaCO_3$ alkalinity added/mg Na_2CO_3 (soda) added

Phosphorus Removal   2.7.2

- Metal Salts
  Typically need about 1.25–1.75 moles of metal/mole phosphorus remaining after background removals to achieve low soluble phosphorus residuals of less than 1 mg/L. This results in the following general representations:
  Alum (0.87 mg Al/mg P at 1 mole/mole):

– 5.6 mg $CaCO_3$ alkalinity loss/mg Al added

– 3.9 mg $AlPO_4$ formed/mg P removed

– 2.9 mg $Al(OH)_3$ formed/mg Al added

≈ 6.1 mg $Al_2PO_4(OH)_3$ formed/mg P removed (assuming 80% P removal) at 1.5 mole Al/mole P remaining for chemical removal

Ferric Chloride (1.8 mg Fe/mg P at 1 mole/mole):

– 2.7 mg $CaCO_3$ alkalinity loss/mg Fe added

– 4.9 mg $FePO_4$ formed/mg P removed

– 1.9 mg $Fe(OH)_3$ formed/mg Fe added

≈ 7.9 mg $Fe_2PO_4(OH)_3$ formed/mg P removed (assuming 80% P removal) at 1.5 mole Fe/mole P remaining for chemical removal

• Lime

– 5.4 mg $Ca_5OH(PO_4)_3$/mg P removed plus

– Precipitated $CaCO_3$, which depends on lime dose and final pH (which will cause the precipitation of $mg(OH)_2$ at pH ≥ 11); lime dose is influenced by wastewater alkalinity

• Enhanced Biological Activity

– 3.6 mg cellular storage products/mg P removed beyond normal background conditions

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of external additives is included in the tabulation of information.

A listing of phosphorus removal reactions is also provided in Table 2-3 since current practice often finds treatment standards that specifically limit both nitrogen and phosphorus. As shown, phosphorus removal can be achieved with a variety of well-demonstrated approaches. The metal salts that are routinely added for phosphorus removal also cause an alkalinity depletion. The metal salt alkalinity depletion associated with phosphorus removal (or if such depletion occurs as a result of other uses, such as dewatering) must be considered when designing for control of nitrogen and phosphorus.

2.2.4 Process Design Review

Table 2-4 summarizes the three most important review checks for nitrogen control facilities. The first two checks apply to any wastewater treatment plant design. The last check specifically emphasizes the details of process design with a progressively narrower focus on nitrogen control technologies. The basis for these recommendations flows from the subsequent discussions contained in this chapter, and in the remaining chapters of this manual.

The process design details described in Table 2-4 are not exhaustive and should be used only as general guidance. The guidance is best applied as "flags" for further exploration; other considerations may prove equally important in the successful design of any nitrogen control facility.

2.3 Fundamental Treatment Technology Concepts

The purpose of this section is to introduce the reader to the fundamentals of wastewater treatment, the multiplicity of options, and the generally available body of knowledge for nitrogen control at the time of this manual's preparation.

2.3.1 Wastewater Treatment

2.3.1.1 Overview

In municipal wastewater treatment, the designer and owner should lean toward low-maintenance, tolerant and ample capacity facilities to reflect the uncertainty of staffing, maintenance, and remedial action in a public marketplace where funding of major capital improvements is uncertain and achieved only by public indebtedness with political and public oversight. Both must walk a careful line between providing tolerant facilities that can respond to a multitude of future uncertainties and a blatant overdesign that results in the misuse of public monies for clearly superfluous facilities. When in doubt, trust experience, which strongly suggests that simplicity and harmony with naturally occurring reactions are likely to serve better than a multitude of unit operations for an optimized desktop objective and/or the temptations of an unproven form of high technology.

With the exception of disinfection, all wastewater treatment processes are directed toward separating pollutants into innocuous gaseous or concentrated end products from the liquid flow stream. Each unit process or unit

Table 2-4. The Three Most Important Review Checks for Nitrogen Control Facilities

1. RESIDUALS MANAGEMENT: *If You Remember Nothing Else, Remember This*

- Does the facility have a reliable means of solids handling and disposal, conservatively designed to meet the actual needs of the intended operation, with some backup should the preferred mode of operation go down or fail to be realized as anticipated?

2. FUNDAMENTAL FLAWS: *If the Assumptions Are Wrong, the Conclusions Probably Are Too*

- Do the per capita wastewater characterizations make sense for the service area?
- Were maximum and minimum design wastewater characteristics logically defined and used in the design, and are they consistent with the compliance interval defined in the plant's discharge limitations?
- Can the performance of the plant meet the controlling compliance interval and pollutant restriction; is the anticipated average day performance value of the technology at least one-half of the allowable monthly maximum?

3. PROCESS DESIGN DETAILS: *Last, but Often Not the Least*

- Mass Balance
  - Was a mass balance for all pollutants of concern prepared?
- Waste Solids
  - Does the design fairly anticipate the waste solids gain due to the application of external additives and the recycles from the solids processing train?
- Waste Solids Recycles
  - Does the design anticipate the recycles as they will be experienced?
  - How were the soluble nitrogen recycles addressed?
- Systems Intended to Maximize Natural Biological Denitrification and Enhanced Biological Phosphorus Removal
  - Has the design been checked for weaker wastewaters than anticipated in the basis of design?
  - Does it have backup strategies, embodied by external additions (substrate and metal salt), should the system not perform as anticipated?
- Final Sedimentation with Effluent Filtration
  - Is shrinking the final sedimentation system because of a following effluent filter likely to yield an effluent filter that cannot respond to the high flow conditions and needs?
- Alkalinity Check
  - Were the nitrification, metal salt addition, and other chemical demands determined?
  - Was credit taken for alkalinity return derived from planned biological denitrification?
  - Are external sources of alkalinity needed?
  - What about the alkalinity demand for the discontinuous recycle of ammonium-laden returns back to the main flow stream?
- External Additions of Foreign Substrates (e.g., methanol) for Denitrification
  - Will the biological system automatically respond to the addition of foreign substrates (e.g., nonfood wastes such as methanol)? The acclimation response is both compound form and mass specific. Unsatisfactory denitrification performance, even with proper driving substrate additions, and foreign substrate pass-through will occur until the acclimation response is complete.
- Suspended Growth Nitrogen Control System Specifics
  - Is the solids residence time, θ_c, of the nitrogen oxidation culture approximately 7 to 10 days when wastewater temperatures are 15°C (59°F) or lower; is the CBOD$_5$ loading 0.1–0.15 g CBOD$_5$/g MLVSS/d?
  - Is the design MLSS or the return sludge concentration too high for the anticipated flow regimes and sedimentation system?
    MLSS and return sludge concentrations, respectively, of 2,000 and 7,000 mg/L are likely safe under all flow regimes for surface overflow rates (SORs) of up to 70 m³/m²/d (1,700 gpd/sq ft); values 3,000 and 10,000 mg/L respectively, may be unsafe under elevated flow regimes for most sedimentation system designs with SORs of 50 m³/m²/d (1,200 gpd/sq ft) or more).
  - Is the return rate consistent with the return sludge concentration under the flow regime of concern (e.g., the maximum flow week or day)?
  - Has a means for bulking sludge control been provided?
    Return sludge chlorination is safest even with selector technology.

Table 2-4. The Three Most Important Review Checks for Nitrogen Control Facilities (continued)

3. PROCESS DESIGN DETAILS: *Last, but Often Not the Least* (continued)

– Is the oxygen transfer assumption appropriate?

Using one-half of the clean water transfer rate for design is generally protective unless dealing with fine pore diffuser designs at the head end of a plug flow reactor; 75 percent of the clean water transfer rate may be unrealistically high for design purposes without specific justification.

– Has the oxygen supply been checked under all carbonaceous and nitrogenous loading conditions and does it allow the plant to respond to all reasonably anticipated conditions throughout its design life?

– Are design nitrification kinetic rates based on the correct percentage of nitrifiers in the MLVSS?

The number of nitrifiers is determined by the available nitrogen which changes less than the applied carbon with broad applications of treatment technology. The biomass formed by the carbonaceous substrate acts to dilute the number of nitrifiers with increasing carbonaceous loadings. This causes changes in nitrification rates when measured per unit of total MLVSS as discussed in Section 3.3.6. Use of the solids residence time design procedure (Section 6.4.1.1) is an alternate approach.

– For seasonal permits, is the month before the coldest temperature month that the nitrification effluent standard is applicable also considered when selecting the design temperature?

Permit conditions must be satisfied for the entire month. This can only be ensured by designs that allow for attainment of the operating conditions for the required level of nitrification in the month that precedes the month that the nitrogen control effluent standards apply.

– Are the denitrification rates and extent of denitrification reasonable?

High-level denitrification can only be ensured with ample available soluble substrate. The reduction of 1 mg NO_3^--N will satisfy a wastewater oxygenation requirement of 2.9 mg COD. A conservative denitrification design would assume that only 50 percent of the applied soluble substrate is directly available for the denitrification reaction; the balance of the denitrification reaction achieved is by bacterial stabilization of trapped applied wastewater particulates and cellular respiration.

The endogenous denitrification rate can be no greater than the equivalency of a reasonably anticipated endogenous solids destruction or respiration rate. If the overall endogenous solids destruction rate is 0.02–0.10 g VSS/g MLVSS/d, the matching stoichiometric denitrification rate is 0.01–0.05 g NO_3^--N reduced/g MLVSS/d. Wuhrman and others have shown that the endogenous oxygen consumption rate with nitrate as the electron donor is about one-half of that using oxygen; so in actual practice the solids destruction rate of 0.02–0.1 g VSS/g MLVSS/d would normally correspond to observed rates of about 0.005–0.03 g NO_3^--N reduced/g MLVSS/d.

- Attached Growth Nitrogen Control System Specifics

– Were design strategies for "as required" operator implementation provided to counter unique SS control issues for trickling filter and rotating biological contactors?

Sloughing should be considered. Means of downstream capture should be considered. Many studies have shown that in separate-stage nitrifying trickling filters the effluent SS equal the influent SS; therefore, permit requirements will dictate the need for downstream clarification (or filtration) in these cases.

Attached growth systems produce a difficult to capture colloidal suspension. Use of suspended growth solids contactor concepts with combined carbon and nitrogen oxidation technologies, with design MLSS concentrations of 500–1,500 mg/L, will produce a more visually attractive, lower SS effluent quality.

– Is the nitrification technology correctly applied?

Municipal wastewaters have trash that can clog the media. Upstream clarification or fine screening is usually necessary to avoid media blinding and clogging.

– Is the target NH_4^+-N concentration reliably achievable?

Unlike suspended growth systems, attached growth system nitrification performance may be more limited by oxygen transfer than temperature. Additionally, nitrification technology with attached growth systems in either a combined or tertiary application becomes progressively uncertain as the desired effluent ammonium levels become very stringent.

– Does the proposed nitrification system have the needed auxiliaries?

A recirculation capability should be provided for all attached growth systems unless there are consistent provisions for low BOD in the applied wastewater or other design provisions are made. Recirculation inherently sends nitrifiers to the front of the system, provides favorable dilution of soluble biodegradable carbonaceous compounds, and is the means of providing more oxygen to the applied load for trickling filters.

Trickling filters should have the ability to control the instantaneous application rate independently of flow. Care should be exercised in the election of some cross-flow media with combined carbon and nitrogen oxidation systems. Vertical flow media is a safer design choice if there are clogging concerns.

3. PROCESS DESIGN DETAILS: *Last, but Often Not the Least* (continued)

Rotating biological contactors should have the ability to periodically clean (air scour) the carbonaceous stages of excess biogrowth and remove settled solids from the bottom of each stage.

Tertiary applications of either technology may benefit from the ability to receive upstream diversions of clarified raw or partially stabilized wastewaters.

— Is the denitrification system correctly sized for its application?

Successful denitrification in tertiary applications requires bacterial acclimation to the foreign substrate and sufficient organism numbers to achieve the desire temperature-dependent reaction in the available residence time. A dirty bed operation is mandatory when denitrification is incorporated into effluent filtration, and hydraulic application rates may be appreciably lower than routinely encountered with conventional effluent filtration applications. Overall system performance will be influenced by the time in service of all parallel reactors. Limiting the number of parallel reactors may significantly compromise overall performance as units are removed for cleaning and acclimate on return to service. The need for periodic rigorous cleaning must be anticipated.

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operation is preparatory for the next. Those processes that result in a gaseous product are normally followed by a subsequent solids concentration step until the liquid or solids residual is suitable for return to the environment. "Everything has to go somewhere."

Acceptable ultimate release to the environment is by one of two ways—dilution or concentration. Management strategies and treatment technologies that promise the maximum dilution are likely to be superior to those that promise the maximum concentration, and should be selected first if the opportunity presents itself. Similarly, treatment technologies with the fewest moving parts are likely to be superior to those with many and should be selected first when the opportunity presents itself.

The foregoing paragraphs provide a generalized set of ideals that can be used to guide project decision-making. Practicalities, of course, ultimately govern. Project decision-making must ultimately balance the human and environmental resources of time, space, energy, and capital. Indeed, most key project decision-making becomes more a matter of "what you can't do" than "what you can do," and some unit process selections become the only option available for achieving compliance with the needs of the regulator and public in the time available for implementation.

2.3.1.2 Technologies

Table 2-5 provides a categorical description of the processes typically encountered in municipal wastewater treatment. The intent of this discussion is twofold. First, it is to establish the fundamentals of wastewater treatment from several broadly based perspectives that may well serve the designer and reviewer in their consideration of the overall process scheme and specific unit process options. Second, it is to provide the reminder that overly optimistic design assumptions elsewhere in the process design may doom to failure even the most conservatively designed nitrogen control facility within the integrated works of the treatment facility. For more infor-

mation than is provided here, the reader is referred to References 1 and 3.

A wide variety of unit processes and processing options are available to the designer. Procedurally, they consist of a series of reactors and separators. Reactors create gaseous end products, oxidize, reduce, solubilize, immobilize, or physically condition. Separators result in low and high solids product streams. Reactor and separator processes can be "passive" or "reactive," with their design and performance generally influenced by some key dependency on hydraulics, pollutant concentration, or pollutant mass. The majority of the unit processes are passive and largely sized on hydraulic considerations.

The performance of "passive" unit processes is not strongly subject to operational manipulation. Here, the designer has the responsibility to err on the conservative side, since "not enough" would mean "provide more" because few, if any, remedial operational strategies are available to counteract the undersized units. Well-designed, passive unit processes are preferred for plant owners/operators with a low personnel commitment and a desire to avoid frequent attention. Such plants largely run themselves or, at least, require less operator attention. They are often the most expensive in capital investment and/or expansive in areal commitment. Some are unresponsive to new or unanticipated treatment requirements, and, once upset, may take the longest to recover.

The performance of an "active" unit process is the opposite of the passive process. It is easier to upset, and to turn around from an upset condition, than in a passive process. The simplest active process is a chemical mix tank receiving some additive. Process "activity" is also promoted by the provision of some recycle to manipulate the responsiveness of the reactor. Active processes allow field optimization and usually present an opportunity to derive first and operating cost savings. Some are attractive for seasonal operation, or when flows and/or loads

Table 2-5. Classification of Wastewater Treatment Processes

Processing Objective	Representative Technologies	Process Format[a]	Key Design Criteria
• Preliminary Treatment—Solids separation and conditioning for easier downstream treatment	Screening and degritting separation	Passive	Hydraulics
• Primary Treatment—Solids removal for more economical downstream processing	Sedimentation tank separation	Passive	Hydraulics
• Biological Treatment—Solids and oxygen demanding substrate removal by oxidation, synthesis, and conditioning	Suspended and/or attached growth reactor	Passive to active	Varies per technology: hydraulics, pollutant concentration and mass, solids residence time, returns, operating solids levels, character and specific surface of media, and nature and variability of wastewater
• Disinfection and Dechlorination—Control infectious agents and residual chlorine toxicant	Chlorination oxidation and sulfite reduction, ozonation	Active	Hydraulics, dose, and mixing
	UV radiation	Passive to in-between	Hydraulics and dose
• Integrated Processes for Enhanced Biological Nutrient Control—Use reducing anoxic processing environments for N and P removal beyond background synthesis values	See Biological Treatment: multistage processes can be mainstream, sidestream, and incorporated offstream; suspended growth process mandatory for present understandings of biologically enhanced phosphorus removal; high-level nitrogen elimination may only be ensured with supplemental nitrogen-free substrate additions (e.g., methanol)		
• Add-on Processes for Advanced Waste Treatment—Removal of P, N, SS, soluble biologically resistant organics, and dissolved inorganic solids from liquid processing train	Depends on application: most common	Passive Active In-between	Hydraulics and SS Dose Dose and hydraulics
	Nitrogen Control—see Biological Treatment		
	SS—filtration		
	P control—Fe and Al salts; lime		
• Sludge Conditioning—Facilitate subsequent handling or ultimate disposal	Chemical addition and/or grinding	Active Passive	Solids mass Hydraulics
• Thickening and Dewatering Separator/Concentrators—Facilitate subsequent processing and/or ultimate disposal	Gravity thickening	Passive	Hydraulics
	Beds	Passive	Hydraulics and seasonal exposure
	Other thickening and dewatering	In-between to active	Hydraulics, solids mass, flotation air supply and pressure, and chemicals
• Sludge Storage—Facilitate subsequent processing or ultimate disposal	Tanks or piles	Passive	Hydraulics and solids concentration
• Sludge Stabilization—Further conditioning to control a sludge's putrescible fraction and other beneficial results	Chemical addition (see above)	Passive to active	Hydraulics
	Digestion and physical pasteurization[b]		
	Composting and other autothermal processes[b]	In-between	Solids mass, biodegradability and concentration

Table 2-5. Classification of Wastewater Treatment Processes (continued)

Processing Objective	Representative Technologies	Process Format[a]	Key Design Criteria
• Thermal Processing—Facilitate downstream processing or disposal by solids conditioning, water evaporation, and/or destruction	Thermal conditioning (and oxidation)[b]	Passive	Hydraulics (and solids concentration and volatility)
	Drying	In-between	Hydraulics and solids concentration
	Incineration	In-between to active	Same as above and volatile solids
• Ultimate Disposal—Environmentally sound reuse and/or disposal	Landfill	Passive	Hydraulics, solids mass and character
	Beneficial reuse	Active	Same as above and reuse strategy

[a] "Passive" process performance is not strongly dependent on operator manipulation. "Active" process performance is strongly dependent on operator manipulation.

[b] These processes necessarily solubilize pollutants in the course of their reaction and, to varying degrees, create soluble pollutant recycles that should be anticipated in the liquid processing stream. The impact of these recycles can be especially pronounced if sludge processing is discontinuous and mainstream processes have low hydraulic detention times and/or depend on, or provide, plug flow reactor configurations.

are expected to change markedly over the effective life of the project.

A process's activity should not be confused with its reliability. The most demanding (and least reliable) unit processes are those that have many support systems and moving parts, and operate under elevated temperatures and/or pressures. Most often, these processes are found in the solids processing train of the plant.

Special care is required for the subsequent solids handling processes where large volumes of SS are encountered; examples are a mainstream activated sludge system or sludge digester. Wasting solids from the liquid processing train is the principal means that an operator has for responding to an effluent that is too high in SS. Poor settling and/or excessive recycles may dictate the sudden need for elevated wastage rates of low-SS concentration streams from the liquid processing train. Overly optimistic settled digested solids concentrations and/or separated supernatant quality design assumptions have resulted in disastrous consequences at many wastewater treatment installations.

The designer will be best served by remembering that the purpose of the mainstream clarifiers is to clarify, not thicken; separate waste solids thickening should be provided prior to subsequent processing. Thickening processes that are most tolerant to widely varying flows and solids concentrations (e.g., gravity and flotation) may be preferable to those that are not (centrifugation and belt). If a gravity thickener is elected, it should be remembered that overflow SS levels can change rapidly from acceptable to disastrous if unrealistic storage needs have also been imposed on the thickener without attention to maintenance of adequate liquid levels above the clarified liquid/thickened sludge interface.

The anaerobic or aerobic digestion process contains an appreciable reservoir of SS. The designer should remember that the ultimate success of the plant depends on the ability to remove sludge permanently from the facility. A single or undersized sludge dewatering unit has a high probability of failure, as does a residuals management plan without a firm outlet. The safest design approach is to assume that the digester is down for cleaning or repairs or that it may not get its anticipated degree of stabilization or underflow concentration when sizing the plant's dewatering or downstream systems. A capability to add lime to the plant's final sludge product ensures greater flexibility and confidence in responding to newly promulgated sludge management regulations and their pathogen, vector, and nuisance control requirements (4,5).

2.3.2 Nitrogen Control Technologies

Table 2-6 summarizes the status of the nitrogen control alternatives in municipal wastewater applications after some 20 years of practice in the United States. The first efforts saw an equal interest in biological and physical/chemical approaches. The fundamental difference between the two is that the former can be designed for either nitrogen oxidation or removal, whereas the latter only provides for nitrogen removal. More subtle differences between the two are found with biological processes offering the conversion of biodegradable organic nitrogen to ammonium, while the physical/chemical approaches leave the dissolved organic nitrogen essentially untouched.

The original biological design approaches were conservative because of uncertain kinetics. These early designs emphasized the better understood suspended growth technologies and often elected isolated cultures (staged activated sludge systems) for the specific processing objectives of carbon oxidation, nitrogen oxidation, and nitrogen reduction. The physical/chemical approaches were

Table 2-6. 1990 Status of Nitrogen Control Technologies in Municipal Wastewater Treatment Applications

	Knowledge[a]			
	High			Low
Technology	Well Demonstrated	Limited Application[b]	Found Lacking[c]	Emerging[d]
BIOLOGICAL TREATMENT				
Higher Technology, Mechanical Plant Approach (see Chapters 6 through 8)				
Suspended Growth				
Single Sludge				
Multiphased				
Aerator and/or Aeration Basin Cycling	O,R			
Sequential Batch Reactor	O			R
Multistaged (e.g., serial application of processes)	O,R			
Multizone (e.g., ditches)	O,R			
Two Sludge	O,R			
Three Sludge	O,R			
Attached Growth, Single- or Multiphased, and/or Staged Applications				
Submerged Media				
Fluidized Bed		O,R		
Packed Bed				
Downflow	R	R		O
Upflow		O,R		
Nonsubmerged Media				
Stationary (e.g., trickling filter)	O	O		R
Rotating (e.g., rotating biological contactor)	O	O		R
Combination Processes				
Any of the Above in Serial Application	O,R	O,R		O,R
Submerged Stationary Media (Vertical Plates or Media)				O,R
Nonsubmerged				
Stationary Media with Solids Recycle (e.g., Activated Biofilter)		O		O,R
Rotating Media in Solids Suspension		O,R		O,R
Specific Surface Additives to Suspended Growth System				
Concurrent Additive Management (e.g., powdered activated carbon)		O		O,R
Separate Additive Separation, Processing, and Return (e.g., Linpor, Captor)				O,R
Lower Technology, Transitional and Natural Systems Approach (see Section 2.4 for more detail)				
Transitional				
Aerated Lagoons (suspended growth)	O	O		R
Intermittent and/or Recirculating Sand Filtration (attached growth)	O	O,R		O,R
Aquatic-Based				
Lagoons (suspended growth)		O,R		O,R
Facultative (N stripping)				R
Algae Harvesting (N removal by stripping & synthesis)[e]		R		
Natural and Constructed Wetlands (attached growth with N removal by synthesis)[e]				

35

Table 2-6. 1990 Status of Nitrogen Control Technologies in Municipal Wastewater Treatment Applications (continued)

	Knowledge[a]			
	High			Low
	Well Demon-strated	Limited Appli-cation[b]	Found Lacking[c]	Emerg-ing[d]
BIOLOGICAL TREATMENT (continued)				
Surface Flow (floating and rooted aquatic plants)	R		O,R[e]	O
Subsurface Flow (with rooted aquatic plants)	R		O,R[e]	
Land-Based (attached growth treatment with and without N removal by synthesis)[e]				
Slow Rate Infiltration	O,R			
Rapid Rate Infiltration	O,R			
Overland Flow	O	R		
Subsurface Infiltration	O	R		
PHYSICAL/CHEMICAL TREATMENT (see Section 2.5 for more detail)				
Ion Specific				
Ammonia Stripping (NH_3-N)		R		
Ion Exchange (NH_4^+-N) (NO_3^--N)		R		
Breakpoint Chlorination (NH_4^+-N)		R		
Non-Ion Specific				
Reverse Osmosis		R		
AVOIDANCE OF NITROGEN CONTROL TECHNOLOGIES THROUGH BENEFICIAL REUSE (not relevant to this manual)				
Irrigation	R[e]			
Selected Industrial Reuse (e.g., cooling water)		R		

[a] O = Nitrogen oxidation; R = Nitrogen removal by biological denitrification unless otherwise noted. Classification can vary depending on particular application.

[b] Knowledge of performance capabilities is high but process has been used only on a limited basis.

[c] Knowledge of performance capabilities is high but process capabilities or economics have been found to be poor, based on limited application.

[d] Knowledge of process performance capabilities is low because of infrequent or recently emerging application.

[e] All systems that rely on synthesis for nitrogen removal ultimately must plan for harvest and disposal of the resultant biomass (10–20 times the synthesized nitrogen is a likely rule of thumb).

attractive to some early designers because of low energy, fear of industrial wastes toxicity, and the determination that the reaction kinetics should be more certain.

Practice soon revealed that the physical/chemical processes had a multitude of problems in full-scale application. Some problems were associated with the absence of knowledge about how to design successfully the physical features of the desired works. Other problems were associated with side reactions and precipitates with the wastewater's dissolved and particulate pollutants and the unforgiving nature of rapidly occurring physical/chemical reactions. The physical/chemical processes soon fell into disfavor because of high operating costs, the need for highly skilled operation, and frequent and expensive maintenance, coupled with greater understanding, accep-

tance, and demonstration of the potential biological approaches.

Today, for all practical purposes, physical/chemical nitrogen control approaches have been abandoned for municipal wastewater treatment except for polishing applications, where further nitrogen control may be a particular design objective (leading to the selection of ion exchange) or is a natural result of the use of a non-ion-specific technology (e.g., reverse osmosis).

The application of attached growth technologies for nitrogen control soon followed the suspended growth systems. Concurrently, and progressively thereafter, greater understanding of the suspended growth systems and the economic issues associated with culture isolation (sedimentation tanks to serve each culture) led to the integra-

tion of the various cultures and processing objectives into single-sludge systems. Today, single-sludge nitrogen control systems are found in a variety of multipurpose applications. As in the past, the evolution and application of attached growth and combination processes for similar objectives continue to lag the suspended growth systems.

Familiarity with a technology leads to a greater understanding of its proper application. Over the last decade, the field has begun to recognize that election of a higher technology, mechanical plant approach may be a poor choice for smaller, land-rich communities when climatological conditions will allow a lower technology, transitional or natural system approach. Unfortunately, little is known about the variability of performance in terms of the worst month and week per year compliance intervals that are encountered in the permitted effluent standards. It is reasonable to expect that performance variabilities become appreciably higher as greater reliance is placed on the biota and media encountered in a natural or man-made setting that is highly subject to the whims of nature. This consideration leads to the inevitable concern that the risk of noncompliance may be a problem with acceptance of the lower technologies. Additional data are needed to confirm or mitigate this concern. However, even if true, an occasional excursion is far more acceptable than chronic noncompliance through the use of inappropriately complex technologies that cannot be properly operated and maintained under local circumstances.

2.3.3 Processes of Principal Focus for This Manual

The foregoing perspective explains the categorical determination of the knowledge found with the nitrogen control technologies (Table 2-6) and leads to the focus of this manual. The categorization varies for some of the treatment systems depending on the particular application under consideration. The processes of principal interest to this manual are those that are well demonstrated and that have proven successful with the widest possible application in municipal wastewater treatment. These technologies almost exclusively lie in the higher technology, mechanical plant approaches and presently emphasize suspended growth more so than attached growth systems because of natural evolutionary process development.

Table 2-6 indicates that none of the higher technology, mechanical plant approaches have been found to be lacking, with the possible exception of powdered activated carbon (PAC) addition. Here, the problem is only with the wet oxidation regeneration step needed for economical application of the technology (PAC regeneration and reuse). To date, separation of the regenerated activated carbon fully from the naturally occurring background wastewater inerts and the associated economic issues have served to limit the routine acceptance of this technology.

As shown in Table 2-6, questions regarding the universal applicability of some of the attached growth systems are encountered because of the uncertainty of the technology in submerged, packed bed and nonsubmerged configurations when high-level nitrogen removal requirements are imposed in colder climates or low-level effluent ammonium-nitrogen requirements are encountered. This is attributable to their inherent process inflexibility (passivity) and the plug flow nature of their performance. Section 2.6 provides greater fundamental understanding of these constraints, as do Chapters 6 and 7.

2.3.4 Approaches That Receive Limited or No Attention in This Manual

The nitrogen control approaches receiving limited or no attention in this manual are the concepts embodied in lower technology, transitional and natural systems most applicable to small communities; physical/chemical nitrogen control strategies; and the beneficial reuse wastewater management practices that allow avoidance of nitrogen control systems altogether. The following paragraphs provide additional understanding of these concepts.

The lower technology approaches are at the same or lower point in understanding than the higher technology, mechanical biological treatment plants and physical/chemical nitrogen control processes were 20 to 30 years ago. Today, they can be considered only an emerging technology. Broadly based acceptance and understanding of technology limits will come with further investigative knowledge. The only certainty is that they are highly attractive conceptual alternatives for rural and/or land-rich communities that should work, in some fashion and to some performance level, under the applicable constraints imposed by the natural and/or structured changes to the environment. Section 2.4 provides additional detail regarding the lower technology, transitional and natural system approaches to nitrogen control. It is provided more to stimulate, characterize, and offer alternate references for additional investigation, than to offer finite design information.

The physical/chemical processes for nitrogen control are at the opposite end of the spectrum from lower technology approaches. Although receiving only limited application, there is enough knowledge to determine that they have limited or no potential for most municipal applications. Ion exchange (NH_4^+–N type) may be lacking in terms of suitability for the primary N-removal system in municipal wastewater treatment, but may have use as a tertiary polishing unit (NO_3^-–N type) to meet demanding effluent requirements. Ammonia stripping may have limited roughing applications, most attractively applied in warm temperature pond settings. Breakpoint chlorination can only be recommended for backup polishing applications to ensure extremely low or nonexistent ammonium residuals. Reverse osmosis cannot be recommended specifically for

nitrogen control, but invariably achieves its removal by the very nature of the process itself. The physical/chemical processes are briefly discussed in Section 2.5, more in the interest of completeness and to point out the problems of the past in order to avoid their repetition rather than to recommend their use.

Actual avoidance of treatment, although not relevant to this manual, can be a more appropriate nitrogen control alternative for some applications. The owner and designer should explore the opportunities for beneficial reuse and injection at the beginning of the project rather than at its end, or not at all. The City of St. Petersburg, Florida, for example, has successfully incorporated such strategies (i.e., cooling water reuse and irrigation for recreational areas, with the balance injected) for all of its 2,200–2,400 L/s (50–55 mgd) of wastewater, while minimizing the need for further development of raw potable water supplies and avoiding expensive nutrient removal wastewater treatment strategies at its four wastewater treatment plants. Similar approaches are under active consideration by other Florida communities facing stringent total nitrogen effluent limits (equal to or less than 3 mg/L). These approaches promise to be more frequently encountered at coastal and/or arid regions of the country.

2.4 Lower Technology, Transitional, and Natural System Approaches

This chapter opens with the results of a survey of 150 small plants with debilitating problems (Table 2-1). The number one problem has been poor operator understanding and application of process control. The number one conclusion has been that activated sludge (suspended growth) treatment may not be a good choice for many small plants and the consequent recommendation was to give fuller consideration to simpler, more tolerant treatment processes (e.g., fixed media and natural systems) that are less dependent on skilled operators. Undoubtedly, in some manner, the number one problem was linked to the second cited problem area, solids handling and disposal (too high an operating solids concentration for the elevated flow regimes associated with the number three problem, infiltration and inflow, or "couldn't" or "didn't" remove sufficient solids from the plant). Regardless, the conclusion is inevitable. When a choice can be made, beware of the temptations of elaborate technology.

In the preceding section, lower technology approaches for nitrogen control were identified as emerging technologies. In areas rich with land, their acceptance has been hampered by one or more of the following: failure even to consider (business as usual); institutional reluctance (it is more convenient and often more economical to procure a packaged system that promises to meet needs; sound design of any kind takes time and uses skilled, expensive people); zealous single-issue advocates (anything natural is better, resulting in failure related to inappropriate appli-

cations); imperfect perceptions (not in my backyard); historic practices (the failing septic tank, honey wagons and farms, the stinking raw sewage lagoon); implementability (the required acquisition of large tracts of land); technical uncertainty (design criteria, performance variability, clearly incomplete reporting of costs and operating issues); and potentially overly restrictive discharge standards (which impose unacceptable noncompliance risk).

Lower technology approaches are not new, nor are they uncommon (1). Even today more than 50 million U.S. residences (25 percent of all single-family dwellings) have remained unsewered, mostly served by septic tanks or soil absorption systems for their wastes. Stabilization ponds (or lagoons) number well over 5,000. Specifically designed land treatment systems number over 1,000. Constructed wetlands number over 100, along with a few aquatic plant systems. Nor are land-based systems always small—Orlando, Florida, uses a rapid infiltration land treatment system (ground-water recharge) with a 2,200-L/s (50-mgd) design capacity.

Serious consideration should be given to lower technology, natural system approaches. Advocates of both higher and lower forms of technologies should be open to each other. The best approach for a midsize, land-rich municipal system may not be one or the other, but may be a blend of both.

The purpose of this section is to introduce the reader to this technology as it applies to nitrogen control applications. The balance of the section is devoted to providing additional understanding of the fundamentals and of the treatment technology found in broadly based generic classifications. Thereafter, additional understanding is best derived from the many readily available publications that consider this technology in a summarial fashion (1), comprehensively (6,7), and with technology-specific detail (8–12). The *Manual: Wastewater Treatment/Disposal for Small Communities* should be consulted for guidance in any application specifically dealing with a limited population base (13).

2.4.1 Fundamentals

Table 2-7 compares both the higher and lower forms of technology using readily understandable screening criteria: the land required for the process and the ability of the technology to meet the nitrogen control objectives. The table also includes three transitional technologies that effectively blend the mechanical plant with the natural system approach.

An inspection of Table 2-7 shows the attractiveness of the higher technology, mechanical plant approach: significantly less land required and the ability to somewhat easily respond to all of the target effluent objectives except the most stringent total nitrogen standard. If all of the lower technology, natural system approaches were placed in a similar grouping, the only fair conclusion

Table 2-7. Lower Technology, Transitional, and Natural System Screening Criteria

	Land Need acres/mgd ADF[a]		Detention Time	Nitrogen Oxidation, Effluent NH₄-N			Nitrogen Removal, Effluent Total N		
				Estimated General Applicability of Technology for Various Levels of Nitrogen Control					
Technology	Site[b]	Process[c]	Days	Inter-mediate (5–10 mg/L)	High (2.5–5.0 mg/L)	Stringent (<2.5 mg/L)	Inter-mediate (10–15 mg/L)	High (5–10 mg/L)	Stringent (<5 mg/L)
HIGHER TECHNOLOGY, MECHANICAL PLANT APPROACH,[d] Suspended and Attached Growth Technologies (provided for reference)									
	0.5	0.2 +/-	1.0 +/-	Easy	Easy	Easy	Easy	Easy	Possible
	Varies with technology (see Tables 2-6 and 2-8)								
LOWER TECHNOLOGY, TRANSITIONAL APPROACH,[d] Suspended and Attached Growth Technologies, Oxygen Supply by Mechanical Means or Resting, Respectively. (Note: Imhoff or septic tank preferred for pretreatment for primary settleable solids and floatables capture for the latter two technologies.)									
Aerated Lagoons	—	3–30	10–100	Possible	Possible	Doubtful	Possible	Possible	Doubtful
Intermittent Sand Filtration	—	2–10	—	Easy	Possible	Doubtful	Doubtful	Doubtful	Doubtful
Recirculating Sand Filtration with Anoxic Rock Filter Enhancement	—	8–10	1–2	Easy	Possible	Doubtful	Easy	Possible	Doubtful
LOWER TECHNOLOGY, NATURAL SYSTEMS APPROACH[d,e] (Note: Imhoff or septic tank preferred for pretreatment for primary settleable solids and floatables capture for submerged flow wetlands and subsurface infiltration; lagoons for other land-based systems.)									
Aquatic-Based Technologies									
Lagoons, Suspended Growth Technology with N Removal by Stripping (small), Nitrification and Denitrification, and/or Synthesis; Oxygen Supply Largely by Photosynthesis[e]									
Facultative	—	4–20	20–100	Easy	Possible	Possible	Possible	Possible	Doubtful
Algal Harvesting	—	10+	10+	Easy	Easy	Possible	Easy	Possible	Doubtful
Constructed Wetlands, Attached Growth Technology with N Removal by Synthesis, Oxygen Supply Largely by Photosynthesis[f,g]									
Free Water Surface	—	25–30	7–10	Possible	Possible	Doubtful	Possible	Possible	Doubtful
Submerged Flow	—	5–10	1–2	Doubtful	Doubtful	Doubtful	Doubtful	Doubtful	Doubtful
Land-Based Technologies, Attached Growth Technology with and without Nitrogen Removal by Synthesis, Oxygen Supply by Resting[f,h]									
Slow Rate (Moderate)	—	50–700	—	Easy	Easy	Possible	Easy	Easy	Possible
Rapid Infiltration (High)	—	3–60	—	Easy	Possible	Doubtful	Possible	Possible	Doubtful
Overland Flow (Low)	—	15–110	—	Easy	Possible	Doubtful	Possible	Doubtful	Doubtful
Subsurface Infiltration (Mod - High)		8–80	—	Easy	Possible	Possible	Possible	Possible	Doubtful

[a] Average daily design flow.

[b] Site needs are highly variable. Likely minimum listed for higher technology, mechanical plant approach. Rule of thumb for lower technology, natural system approach would be twice as large as the process requirement, with no site-specific information to the contrary.

[c] Process needs limited to biological reactor and clarifier (if appropriate).

[d] Avoid rigid thinking. Best process selection may be a blend of technologies, both coupled and uncoupled, using each for what it does best. Natural systems are readily adaptable in add-on tertiary, polishing, or seasonal applications.

[e] All systems that rely on synthesis for nitrogen removal must provide for harvest and disposal of the resultant biomass (10–20 times the synthesized nitrogen is a likely rule of thumb).

[f] Readily available information is not clear as to original design criteria (e.g., ac/mgd); generally, existing conditions reported. Additionally, lowest process acreage may be associated with treatment strategies not necessarily directed to nitrogen control. Often systems have not been in operation for several years. Algal harvesting with a view toward maximizing N removal approaches has only been demonstrated at prototype installations in the United States. Some constructed wetlands systems are used in tertiary applications, others provide secondary and tertiary treatment. Often performance data are limited to active growing season.

[g] Inherent anaerobic conditions make subsurface constructed wetland systems attractive for nitrogen removal by denitrification when applied in a staged manner or on receipt of a nitrified feed stream. Nitrogen removal characterizations, if not already noted, rise to "possible" when utilized in this manner.

[h] Typically encountered soil permeabilities are listed parenthetically after each technology. The technologies are generally applied with the following soil cm/hour permeabilities: Slow Rate = >0.15 – <5, Overland Flow = <0.5, Rapid Infiltration = >5.

would be that they too could respond to most of the target effluent objectives, even for the most stringent total nitrogen standard (more stringent standards result in an increase in biomass production, land required, or seasonal restrictions).

Clearly, the lower technology approaches, whether transitional or natural, offer a rich mine of possibilities that have been little explored or optimized to date. With a regulatory and owner attitude willing to accept the risk of the unproven, the opportunities for enhancement and refinement seem almost endless. Apparently all that is needed is a greater appreciation and a sound application of fundamentals. Section 2.6 of this chapter provides discussion of the fundamentals needed for sound application of a lower technology approach: the solids residence time, the carbon-to-nitrogen (C:N) ratio, the differences between suspended and attached growth reactors and their responsiveness, an understanding of oxygen transfer and alkalinity/pH inhibition of the obligate aerobic nitrifying autotrophs, the need for driving substrate for the denitrification reaction, and the mass balance, based on the total oxygen demanding materials (Section 2.9).

To assist in the visualization of community applications of lower technology nitrogen control systems, Table 2-7 also includes a fundamental characterization of the suspended or attached growth type of biological reactor and, because of the importance of the oxygen supply in nitrification (and any desired aerobic reaction), the principal source of oxygen for each system. When the technology relies on synthesis (or plant uptake) or some degree of ammonia stripping for nitrogen removal, this too is listed.

The mass balances for these systems, with the exception of the intermittent and recirculating sand filters, must be expanded to include hydrological considerations (including the wettest and driest year expected in the life of the facility) and, when applicable, the biomass produced by photosynthetic oxygenation (light is the energy source, temperature influences algal activity) and the resultant biomass planned for harvest or accumulation. If accumulation (no harvest) is planned, care must be taken that it does not impair the performance of the system. The wisest course of action for the aquatic systems that use the least amount of land per unit of flow processed, is to plan on the eventual need for vegetative debris (solids) removal and disposal (a reliable residuals management scheme).

More detailed inspection of Table 2-7 shows some aspects of the mechanical plant approach to be more prevalent in lower technology applications. In general, primary treatment is recommended for improved downstream process stability, and freedom from nuisance or operating problems. Undoubtedly, today's failure to accept a lower technology is due in part to the absence of primary treatment in many of the historic applications. Significantly, the success of the downstream process is dependent on what

goes on before it, and no system is completely free of the need for regularly scheduled maintenance and knowledgeable operation.

True natural systems always bring with them valid issues of transient and resident vectors, and their control needs. Control of the human population (via site access barriers) may also be perceived as important and, depending on the technology, it may be. However, before the need for isolation from the human population is accepted as a foregone conclusion, serious consideration should be given to passive recreational activities (e.g., trail use and bird watching) to increase local acceptance and value.

All natural systems that do not make use of artificial barriers (liners) must be evaluated and monitored for ground-water impacts. Ground-water discharge permits are now required in several states and are inevitable in all others. Up- and downgradient ground-water monitoring wells must be included in the design.

Finally, the intrinsic values and liabilities of the lower technology approach beyond the obvious operational and mechanical simplicity and land requirement must be considered. The intrinsic value is the tolerance of time (mitigation of processing flow and pollutant peaks in the applied wastewater and daily attentive operation and maintenance) and dilution (through the commitment of expansive soil and/or liquid volumes for beneficial treatment). The intrinsic liability is the greater unpredictability of relying on some harmony of the processes found in the natural environment with the often unpredictable climatological conditions and media change with time. Approaches to overcoming the liabilities embody the same principles often used in mechanical plants: nonaggressive design, fuller use of recycles or returns, load-splitting capabilities, and low cost backup strategies (flexibility) whenever possible. The remaining subsections address the lower technologies in additional detail.

2.4.2 Transitional Approach

2.4.2.1 Aerated Lagoons

Aerated lagoons have potential in nitrogen oxidation applications. Performance is limited by deposition of viable organisms. Shorter detention times will reduce the impact of suspended algae. The 7- or 10-day detention aerated lagoon is normally not well enough mixed to sustain the nitrifiers, and Parker reports finding no aerated lagoons with a significant amount of nitrification. In the absence of adequate mixing, the nitrifiers must reside on a surface and there is not enough surface to sustain them (14).

If sufficient mixing was provided to keep the viable organisms in suspension, use of aeration systems and/or aeration basin configurations that provide mixing independently of the oxygen supply could provide detention times that should closely approach the same solids resi-

dence time for nitrification as found in the mechanical plants. Although back-mixing undoubtedly occurs, recycle of the aerated effluent would allow a continuous reseed of an acclimated nitrifying culture and may achieve a measure of natural denitrification. When settling is desired, depression traps can be used. If desired, floating sludge collectors can be used to remove the deposited solids. Designers should consider some form of aerated lagoon for intermediate to high nitrification applications in land-rich situations.

2.4.2.2 Intermittent and Recirculating Sand Filtration

Once totaling nearly 500 U.S. installations, the once-through intermittent sand filter was the process of choice for the highest known performance for wastewater treatment plants into the 1950s. It is commonly described in great detail in all regulatory guidelines of that vintage. The 1957 edition of MOP 8 describes the effluent quality as so sparkling clear and odorless that remote discharges had to be posted to warn people against use for drinking (15). Biochemical oxygen demand (BOD) and SS were reportedly routinely below 5 mg/L and, by inference from the reported nitrate data for the technology, the effluent ammonium nitrogen was likely to be routinely less than 10 mg/L.

Recirculating sand filtration was the predecessor of the classical contact bed and trickling filter technologies (the latter adopted out of lack of concern for ammonium nitrogen). The intermittent sand filter predates the rapid infiltration basin of today's land treatment strategies. Similar to a rapid infiltration basin, it was operated in a cyclic manner, with one bed operating, one bed resting, and one bed in cleaning (raking); often a fourth bed was maintained for emergency use under high-flow conditions.

Consisting of 1 m (3 ft) of graded sand and 30 cm (12 in) or less of supportive gravel, the intermittent sand filter was used in raw, secondary, and tertiary applications. Most often it was preceded by an Imhoff tank (still a good decision in order to avoid daily or more frequent removals of primary sludge); operating problems and bed cleaning were clearly minimized with this operation. There is nothing fundamentally wrong with this technology, and it should be rediscovered and considered for intermediate to high nitrogen oxidation applications.

Recirculation provides a means of introducing oxygenated water, nitrate nitrogen, and seed organisms to the applied wastewater. Recirculation rates depend on the objectives and the quality of the wastewater applied to the system. With the recycle of nitrate nitrogen, higher nitrogen removals than achieved with the nonrecirculated system will occur.

The full denitrification potential of the recirculating sand filter system is achieved with exactly the same concepts as typically reported for the higher technology approaches (i.e., the use of a front-end anoxic reactor with nitrified effluent recirculation ratios [R/Q] of 3 to 5, and/or an anoxic reactor receiving the nitrified effluent along with a septic or Imhoff effluent as the internal carbon source to drive the denitrification reaction). In these configurations, the anoxic reactors are typically upflow rock (2–5 cm [1–2 in] in diameter) filters. Reported anoxic reactor empty bed contact times are unchanged for pre- and post-sand filter applications and, based on Q, are 4 to 72 hr (13). The range reflects differences in seasonal site-specific temperatures, applied wastewater and performance needs (presently only described for septic tank effluents), designer certainty, and the need for a highly optimized system. Similarly, the reported physical and operating needs of the sand filter are unchanged in these denitrification applications.

2.4.3 Natural System Approach: Aquatic-Based Technologies

2.4.3.1 Facultative Lagoons and Algal Harvesting

Simple regression-type ammonium and nitrogen removal models of facultative lagoons, have been developed and reported with some suggestion of validation (16,17). These identify pH to be of primary importance, based on an ammonia stripping assumption. A pH rise occurs in the pond because carbon dioxide (CO_2) is the carbon source for the algae, which photosynthetically produce biomass and oxygen. The CO_2 source is largely from the aerobic (surface layers) and anaerobic stabilization (bottom layers and deposits) in the lagoon. With insufficient CO_2, the bicarbonate alkalinity will serve as the CO_2 source, and a significant pH rise can be experienced. Significant ammonia stripping does occurs at a pH of greater than 8.5 (see Section 2.5.2.1 for further understanding).

The reported dependency of ammonia removal on pH could also be a surrogate parameter for an active algal biomass, and the actual ammonium and total nitrogen removals could reflect natural nitrification (using the photosynthetically produced oxygen), denitrification (bacterial use of the dormant algal biomass as the driving substrate during the nighttime hours), and algal synthesis during the daylight hours.

Facultative ponds should be designed to embrace and enhance the anaerobic reactions that produce CO_2 and, most important, methane (CH_4), occurring in the bottom of the pond. Failure to do so will likely result in problems and, inevitably, the progressive buildup of solids and pass-through to the plant effluent. Many past problems with this technology were associated with this consideration. The designer would be well served by consulting the more fundamental publications regarding this technology (18,19).

Facultative ponds have the potential to achieve nitrogen oxidation down to the most stringent levels; their natural daytime to nighttime cycling of photosynthetic activity and aerobic to anoxic bacterial response provides a possible

mechanism of nitrogen removal. Their liability: what to do with the algal biomass once generated. Procedures start with submerged drawoff outlet designs and consideration of chemical coagulation and/or filtration for tertiary algae removal (11). Regulatory standards may allow for a higher effluent SS. Pumped or submerged outlet removal and the sloped sidewalls of the lagoon allow for considerable flow equalization.

Facilities with an algal harvest approach (maximizing nitrogen removal by synthesis) can be designed to incorporate a number of concepts. The large lagoons at Sunnyvale and Stockton, California, return the subsequently removed algae to lagoons with adequate depth to ensure anaerobic activity. The systems have operated since the late seventies with no sludge removal. Alternatively, the pond design could be as shallow as is reasonable and well mixed, with the objective of maximizing light penetration. Algae-removal concepts abound (6,7,11) but are often unused on a sustained basis because of the uncertainty (and now a liability) concerning use or disposal of the harvested algae.

2.4.3.2 Constructed Wetlands

Constructed wetlands are classified as a function of water flow: surface (also termed free water surface, FWS) and subsurface. When simply expressed, constructed wetland treatment technology makes an 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 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, since 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 (20). 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 (20).

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. Unfortunately, application of this approach to nitrogen removal is only in the research stage (13).

Ultimately, the success or failure of the wetland approach for nitrogen control may rest with the harvest of the vegetation (how to remove it and what to do with it), 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. There are no good answers, and the managed protoplasm production systems using constructed wetlands or algal production lagoon systems are the only pure, lower technology, aquatic-based approaches that have the potential to achieve significant levels of ammonium and total nitrogen control on a seasonal and/or year-round basis.

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.

2.4.4 Natural System Approach: Land-Based Technologies

The land-based technologies, although not originally developed for nitrogen control, 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, rather than mere ni-

trogen 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.

The following paragraphs provide additional conceptual information on the four most common land-based technologies. Subsurface, slow, and rapid infiltration systems do not discharge to surface waters and conceptually may allow a more relaxed nitrogen control standard, depending on local ground-water regulations.

2.4.4.1 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.

Subsurface infiltration wastewater management practices are embodied in the horizontal leach fields that routinely serve the more than 20 million septic tanks of individual nonsewered establishments and homeowners. In recent years, they have also been advanced for collective service in small isolated communities.

2.4.4.2 Slow Rate Infiltration

Slow rate land treatment represents the predominant municipal land treatment practice in the United States. The wastewater is applied to a vegetative land surface using sprinkler or surface irrigation techniques and percolates (after losses for evapotranspiration) through the plant/soil matrix. This technology is capable of achieving the highest degree of nitrogen removal.

Typically, assumptions for nitrogen losses due to denitrification (15–25 percent), ammonia volatilization (0–10 percent) and soil immobilization (0–25 percent) supplement the primary nitrogen removal mechanism by the crop. 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.

2.4.4.3 Rapid Rate Infiltration

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.

Designs can be developed that foster only nitrification, or nitrification and denitrification. 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 percent. 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.

2.4.4.4 Overland Flow

Overland flow involves the application of wastewater to the upper section of a gentle, sloping grassland. The thin film of applied wastewater is then collected in runoff ditches for subsequent discharge or further processing. It is designed to be used for relatively impermeable soils or subsoils to avoid infiltration to the ground water.

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.

2.5 Physical /Chemical Treatment Processes

Several physical/chemical nitrogen control treatment processes have been advanced and tried in municipal wastewater treatment applications. Only two remain in routine service. Physical/chemical treatment, except in highly specialized situations, is the process of last resort, especially at small plants.

The reader interested in more detail than provided here is referred to earlier editions of the MOP 8 (21) and of this manual (22). The failures and the reasons for the failures are not well reported. However, these failures stand as vivid testimony to the risk that is involved in out-of-context extrapolation, perhaps superficial interpretation of the results from limited-duration laboratory, pilot, and demonstration studies, or the failure to fully appreciate the attendant routine demands of normal operation and maintenance with an emerging technology. Whatever the reasons, the proof of any concept lies in its applied practice with the people that use it on a daily basis.

The purpose of this section is to give a token introduction to the nitrogen control physical/chemical treatment processes in the hope of providing a theoretical understanding, along with a listing of some of the issues encountered with their application. Some processes remain applicable in practice or in concept; these may be the processes of choice in some municipal applications. However, care is strongly suggested, except in polishing (breakpoint chlorination or ion exchange) or roughing (ammonia stripping) applications.

2.5.1 Fundamentals

When working with physical/chemical treatment technologies, it is important to remember that inorganic ions may be added and background organic and/or inorganic species may enter into the reaction. What happens may or may not be pH dependent; some additives and their reaction may influence the pH. Often, unanticipated severe or not so severe side reactions occur, which only seem obvious with "after the fact" analyses.

Figure 2-1 was prepared to illustrate the pH dependency of several reactions of interest. Some reactions have already been discussed (e.g., the dependency of photosynthesis on CO_2 extraction from the wastewater and its effect on pH with the facultative lagoon concepts discussed in Section 2.4.3.1).

The information presented in Figure 2-1 also includes the pH dependency of hydrogen sulfide (H_2S) dissassociation. As shown, once the pH exceeds 8.3, for all practical purposes it is impossible to have any odors associated with hydrogen sulfide. The pH of 8.3 is of particular interest since it is also the point of equilibrium with the CO_2 in the atmosphere. Given sufficient alkalinity, all liquids, when stripped (or aerated) in any fashion, will approach this value. If the pH is below this value, CO_2 will be removed. If the pH is greater than this value, CO_2 will be added. The overall alkalinity does not change, only its species distribution.

Figure 2-1 will be used, as appropriate, in the subsequent discussions of physical/chemical treatment technologies. The source of the CO_2, carbonate, bicarbonate, and hydroxide relationships are described in a widely available textbook (23). The lime dose and soluble calcium and magnesium distributions in Figure 2-1 were derived for a hypothetical wastewater by use of the theoretical predictions found with Caldwell-Lawrence diagrams (24).

2.5.2 Ion-Specific Technologies

The ion-specific technologies work only on ammonium or ammonia nitrogen. They do not work on organic nitrogen or nitrate and nitrite nitrogen. Raw wastewaters need sufficient bacterial treatment to convert the organic nitrogen to ammonium, but not to a level that would result in nitrite or nitrate production.

2.5.2.1 Ammonia Stripping (NH₃-N)

Ammonia stripping was the conceptual add-on process of choice with lime treatment for phosphorus removal in the early years of physical/chemical nitrogen control technology. Air stripping consists of raising the pH to within the range of 10.5 to 11.5 (see Figure 2-1), a point that achieves very low phosphorus residuals, and of providing sufficient air to strip out the ammonia. The lime (or hydroxide) dosage is very alkalinity dependent (25).

The amount of air needed for stripping can be predicted by the equilibrium relationships expressed with Henry's Law constant. Design airflow rates are typically twice the theoretical prediction. Efficiencies decrease with decreasing temperature.

In the early 1970s, the newly installed South Lake Tahoe Advanced Wastewater Treatment Facility reported ammonia nitrogen removals with its 160 L/s (3.75 mgd), 7.3 m (24 ft) high, packed (redwood media) tower of about 90 percent (effluent NH_4^+-N of 1.4 mg/L). Initial pH values were 10.7; hydraulic loadings were 70 $m^3/m^2/d$ (1,200 gpd/sq ft); the air-to-water volume ratio was about 100:1; and air and water temperatures were about 4 and 12°C (39 and 54°F), respectively (26). The cold weather ammonia-nitrogen removal efficiency declined to 60 percent at an air-to-water ratio of 50:1. Warmer temperature, summer performance was in excess of 90-percent ammonia-nitrogen removal for these loading conditions and a 50:1 air to water ratio.

Ninety-percent ammonia-nitrogen removal at South Tahoe in winter was accompanied by a 8°C wastewater temperature decline through the tower. Cooling towers typically operate with an air-to-liquid volume ratio of 10:1 to 30:1. Thus, to avoid ice formation, operations of stripping towers are effectively limited to wastewater tempera-

44

pH
6   7   8   9   10   11   12

100 ┤   ┌ 0
H₂S, Percent

50   40°C   0°C   50   Ammonium, Percent

0 ┤   └ 100

H₂S and NH₄ Dissassociation

100   HCO₃⁻   100
Percent

50   CO₂   CO₃⁼   50   Percent
                                OH⁻

0   0

pH, CO₂, CO₃⁼, HCO₃⁻, and OH⁻ Relationships

400   400
Mg/L as CaCO₃

Initial Alkalinity = 200 Mg/L as CaCO₃

                    Lime Dose, Soluble Ca⁺⁺ and Mg⁺⁺
200   Ca⁺⁺   200   Mg/L as CaCO₃

      Mg⁺⁺   Ca⁺⁺
                                                Mg⁺⁺
0   0

6   7   8   9   10   11   12
pH
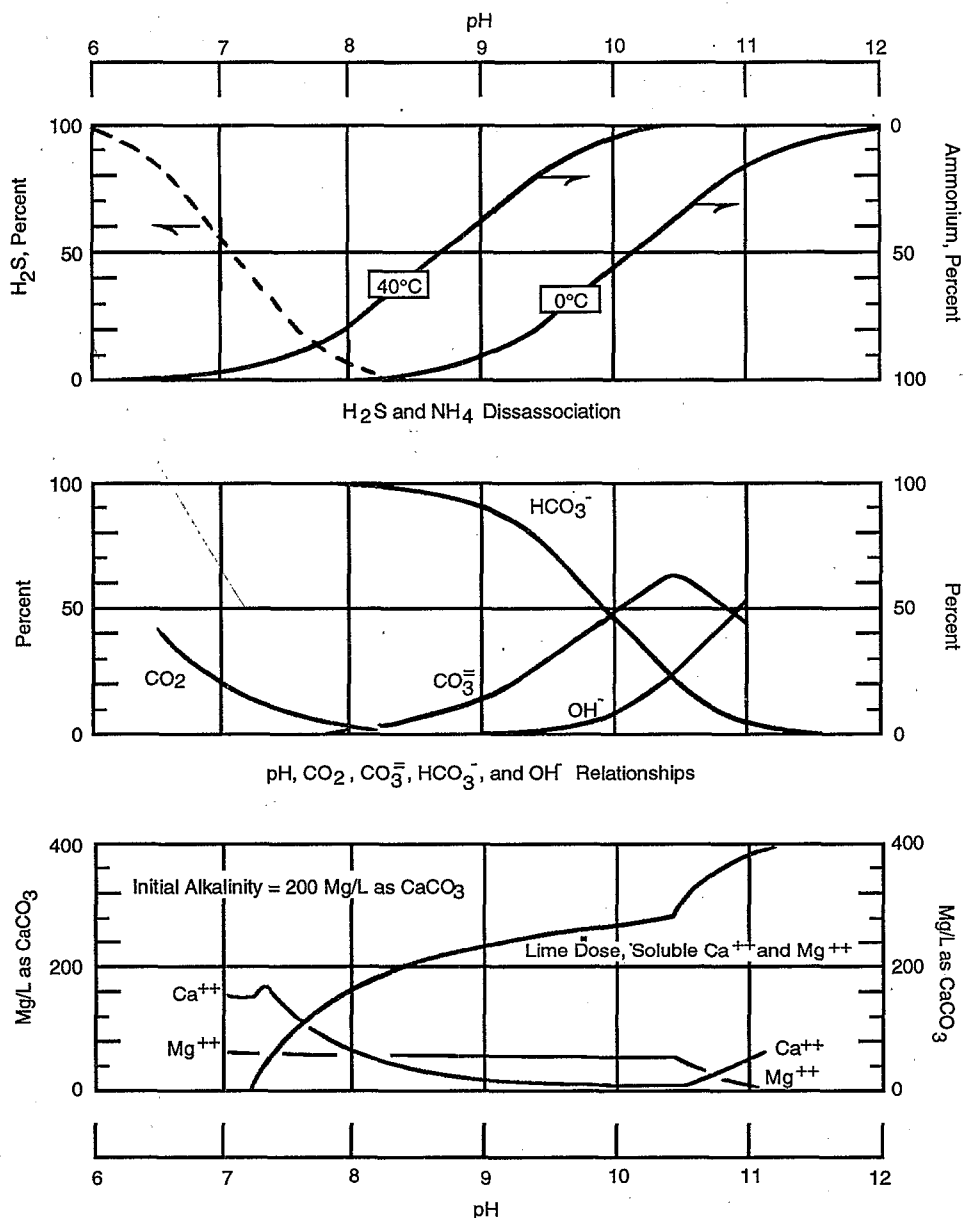
Figure 2-1. pH dependency of selected reactions of interest.

tures of about 10°C (50°F) and above. More important, the 90-percent ammonia-nitrogen removal in winter operation was also accompanied by a 0.4 pH unit decline; this led to the plate-out of CaCO₃ scale, causing serious difficulties.

A survey in the early 1980s of the only two constructed towers found that South Tahoe tried a variety of other packing (e.g., plastic mesh and plastic tubing) and cleaning procedures in the intervening years in an attempt to counter the scaling problems (27). None was successful, all were badly scaled, and, at best the towers could achieve 40-percent ammonia removal (scale formation deteriorated performance). The towers were torn down after seven years, with the conclusion that experience

with no packing, or stripping in ponds, could do just as "good" with much less complexity. Other noted problems were misting (and resultant CaCO₃ deposits) and, with cold weather, ice formation for approximately 60 m (200 ft) around the tower. Other than ice formation, similar findings were encountered with the full-scale stripping tower at Water Factory 21 in Orange County, California. CaCO₃ (calcium carbonate) scale formation problems also continue to plague the downstream pH adjustment step.

The place for ammonia stripping appears to be in warm-climate ponds. It also is used for leachate pretreatment prior to discharge to a POTW. A recent publication reports 50-percent ammonia-nitrogen removal in unaerated ponds in five days with a pH of 10.5, and the same

removal with aerated ponds in about a half day (28). Likely air-to-liquid volume ratios are on the order of 50 to 1. Collection of the treated water and delivery to a following pond, if possible, will achieve the best performance with the least amount of air usage. Surface sprays or mechanical aeration are also applicable.

2.5.2.2 Ion Exchange (NH$_4$-N)

Ion exchange technology involves passing a liquid through a column or bed of specific natural or synthetic zeolite resin and the exchange of one ion for another. The column is run until unacceptable breakthrough of the ion(s) of concern is achieved (reflecting the exhaustion of the exchange sites for this point of equilibrium). A highly concentrated regenerant is then passed through the column to displace the removed ions from the exchange sites. The regenerant can be processed further for recovery and reuse with a blowdown of a concentrated waste or recyclable product, or passed to waste.

Clinoptilolite, a natural occurring product, is the zeolite of choice for ammonium-nitrogen removal applications. Its order of major ion selectivity is reported as (29):

$$K^+ > NH_4^+ > Na^+ > Ca^{++} > Mg^{++}$$
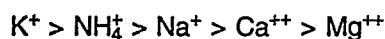
Total exchange capacities are reported around 2 milliequivalents/L (meq/L), but the capacity in wastewater applications for ammonium appears to range from 0.2 to 0.5 meq/L. Clinoptilolite has a specific gravity of 1.6 and a bulk density of about 0.75.

In wastewater treatment applications, prior filtration must be used to avoid blinding the fine 20 x 50 (0.84 x 0.33 mm) mesh media. Hydraulic loading rates range from 5 to 10 bed volumes per hour. On startup, the first 5 to 10 volumes will generally show the dragout of the ammonium left in the column after regeneration. Thereafter, the effluent ammonium nitrogen concentrations slowly increase from about 0.5 to 1.0 mg/L until about 130 bed volumes is reached and then progressively deteriorate to about 5 mg/L at 170 bed volumes. The column may be removed from service after final effluent quality becomes unacceptable.

Regeneration is preceded by two to three bed volumes of backwash to flush the accumulated solids out, and is accomplished by a high pH salt solution using flow rates of about 10 bed volumes per hour and 10 to 20 bed volumes of regenerant. Regeneration is followed by a rinse cycle of 2 to 3 bed volumes to minimize high pH as well as salt and ammonia concentrations in the product effluent on startup.

The two wastewater treatment plants that have clinoptilolite ion exchange use a closed-loop ammonia stripping process to recover the regenerant. The off-gases are absorbed in a sulfuric acid solution to form ammonium sulfate, which has some fertilizer value. In this case, the pH is further adjusted to greater than 11 by caustic, and the resultant Mg (OH)$_2$ is allowed to settle out before entering the ammonia stripping cycle. Makeup salt is added to the regenerant after the ammonia is removed.

The 220-L/s (5 mgd) North Tahoe-Truckee Wastewater Treatment Plant has routinely used its clinoptilolite system since the late seventies. Presently, the plant's average daily flow rate is about 70 percent of its rated capacity.

The 660-L/s (15 mgd) Upper Occoquan Treatment Facility in Virginia, constructed before the North Tahoe-Truckee plant became operational, found operation of its clinoptilolite system to be unnecessary because effluent standards were revised before it began operation. The Upper Occoquan installation was briefly operated for startup and operability checks in the early eighties, found to achieve its 2 mg/L ammonium-nitrogen target, and placed in its present reserve status.

The North Tahoe-Truckee facility historically achieved about 2.5 mg/L of ammonium nitrogen in its ion exchange effluent (because of regenerant dragout and end-of-run deterioration), which was further reduced to about 0.3 mg/L by breakpoint chlorination. The plant's total effluent nitrogen standard was then 2 mg/L. Together, the ion exchange and breakpoint chlorination processes add about 140 mg/L of chloride and 250 mg/L of dissolved solids to the plant's effluent.

Most recently, with relaxation of its treatment needs due to a revised ground-water discharge, North Tahoe-Truckee has discontinued breakpoint chlorination. The plant now averages an ammonium-nitrogen release of 5 to 6 mg/L, with the effluent from any one of the three operating columns allowed to increase to 12 mg/L. This operation is made possible since the columns (four provided) are at various states of ammonium breakthrough (breakthrough occurs rapidly after the operating column effluent exceeds 5 to 6 mg/L ammonium nitrogen).

The view obtained from the North Tahoe-Truckee plant is that clinoptilolite ion exchange is a workable but demanding operation. Maintenance requires frequent hydrochloric acid washing of the closed loop stripper and absorber media to remove the scale. About 20 percent of the clinoptilolite is replaced per year. The caustic, acid, and salt handled in the media regeneration and ammonia recovery process present a corrosive environment, dictating special safety concerns and equipment (e.g., pipes, pumps, valves, fittings, and instruments) which is two to three times more expensive than standard equipment, and difficult to maintain, repair, and replace. However, the plant management is pleased with the process after 12 years of operation.

Operational care is particularly important in terms of preventing ammonium sulfate crystal formation. The formed ammonium sulfate is contaminated with sodium. Once given away to a potato farming operation in Nevada, the

waste ammonium sulfate is now used as a nitrogen supplement in the commercial production of organic fertilizers by others.

The ion exchange discussion above has been limited to large NH_4^+ removal systems. The reader should be aware that a new generation of small (one to a few) home systems is being proposed on a throwaway or commercial regeneration service basis that may have promise. Also, advances in synthetic ion exchange resins may permit their use for polishing of effluents that exceed stringent permit requirements.

2.5.2.3 Breakpoint Chlorination (NH₄-N)

Breakpoint chlorination for nitrogen control originated from the water treatment industry, where it was once used to achieve a free available chlorine residual because of disinfection concerns with chloramines. Figure 2-2 illustrates a typical breakpoint chlorination curve for an advanced treatment effluent.

Figure 2-2 also summarizes the reaction stoichiometry. The reader should come away with the view that many reactions in addition to breakpoint can occur; the process results in the unavoidable introduction of dissolved solids, and it can deplete the alkalinity if a buffer is not added or contained with the chlorine source for this purpose (which may also add dissolved solids). Dechlorination is likely to be necessary for today's municipal wastewater practice and, if activated carbon is not used (which is not recommended because of the expense), this followup process will also add dissolved solids and deplete alkalinity.

The reactions are dependent on pH (around 7 to 7.5 is the preferred target for the fastest reaction rates, while
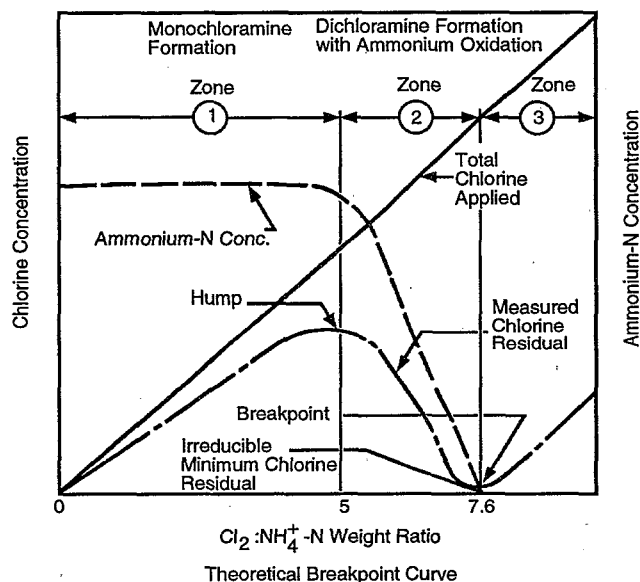
Description	Reaction Stoichiometry*
Breakpoint reaction	$NH_4^+ + 1.5\,HOCl \longrightarrow 0.5\,N_2 + 1.5\,H_2O + 2.5\,H^+ + 1.5\,Cl^-$
NCl₃ formation	$NH_4^+ + 3\,HOCl \longrightarrow NCl_3 + 3\,H_2O + H^+$
Nitrate formation	
1-from ammonia	$NH_4^+ + 4\,HOCl \longrightarrow NO_3^- + H_2O + 6\,H^+ + 4\,Cl^-$
2-from nitrite	$NO_2^- + HOCl \longrightarrow NO_3^- + H^+ + Cl^-$
Other inorganic reactions (e.g., H₂ S)	$H_2S + HOCl \longrightarrow S\downarrow + HCl + H_2O$
Other organic reactions	Organics + HOCl $\longrightarrow$ Oxidation and chlorinated end products

*Percent distribution of HOCl and OCl⁻ with pH

	pH		
	6	7	8
HOCl	97	75	49
OCl⁻	3	25	51

Figure 2-2. Typical breakpoint chlorination curve and stoichiometric reactions for an advanced treatment effluent (from Reference 30).

avoidance of values below 6 is necessary to minimize the formation of toxic, pungent nitrogen trichloride); the chlorine dose; the magnitude of the ammonium nitrogen; and the competition of the various side reactions with nitrite, inorganics (sulfide is presented for reference in Figure 2-2), and organics (trihalomethane residuals are formed once the breakpoint reaction occurs). These reactions affect and extend the breakpoint curve.

Stoichiometric predictions of 7.6 mg of chlorine per mg of ammonium nitrogen are never seen in practice. Even the cleanest processing streams will experience weight-to-weight ratios of 9 or 10:1. Processing streams with high SS and incompletely biologically stabilized soluble residuals can result in weight-to-weight dosages of 20 or more:1.

The mid-seventies breakpoint chlorination experience with a biologically stabilized effluent produced the following end products and chlorine usages (30). The effluent contained a nominal ammonium nitrogen of 20 mg/L; a chlorine to ammonium nitrogen dosage of 10 to 1 was used at a pH point between 7 and 8:

End Products and Chlorine Usages

	Pathway of Applied NH_4^+-N,[a] %	Pathway of Applied Cl_2, %
NH_4^+ to N_2	90	70
NH_4^+ to NO_3^-	5	13
NH_4^+ to NCl_3[b]	2	5
NO_2^- to NO_3^-	—	1
Remaining Residual	3	7
Unaccounted	—	4

[a] Organic nitrogen was reportedly unchanged (likely applied concentration = 0.5 to 1.0 mg/L).
[b] Reverts to NH_4^+-N on dechlorination with sulfite.

The lack of other reported data prevents any conclusion as to the representative nature of the preceding characterization. However, it is likely that 90-percent removal of the applied ammonium nitrogen is achieved once the breakpoint reaction occurs.

The only known operating facility where breakpoint chlorination is the principal nitrogen control strategy is at Sugarbush, Vermont. Here, average daily flows range from about a quarter to one-half of its rated 7 L/s (0.16 mgd) capacity. Ammonium-nitrogen concentrations in the raw sewage reach up to 30 mg/L in the winter, but typically average approximately 15 mg/L. The breakpoint chlorination system is preceded by 7,570 m³ (2 Mgal) of aerated pond storage, alum clarification of the pond effluent, and dual media filters. It is followed by 36,000 m³ (9.5 Mgal) of storage and a 0.4-ha (1-ac) leach field. Chlorine to ammonium-nitrogen weight dosages average 18:1. Nitri-

fication in the aerated raw sewage storage tank during the summer allows a reduction of the chlorine usage, but only with the accompanying result of higher total nitrogen in the processing effluent. The plant flow release is seasonally restricted. Reported November 1990 through October 1991 performance (based on weekly, eight-hour composites) is described below. Note that the average flow was 2.5 L/s (40 gpm) for this period:

Breakpoint Chlorination Performance

	Org-N	NH_4^+-N	NO_2^--N	NO_3^--N	Total N	TRC[c]
Annual Average, mg/L						
Influent	9.3	14.3	—	0.6	24.2+	—
Filter Effluent[a]	2.3	10.5	0.3	1.7	14.8	—
Breakpoint Effluent	—	0.1		3.7	—	6.2
To Field[b]	1.0	0.2		3.8	5.0	—
Worst Month Quality, mg/L						
To Field[b]	1.0	0.1		7.2	8.3	—

[a] 35 days later
[b] $Q = 0.0029 m^3/s$; 140 + 35 = 175 days later
[c] TRC = total residual chlorine

The utilities director's recommendation for others considering full nitrogen control by breakpoint chlorination can be summarized in one word—"don't."

It is recommended that breakpoint chlorination be routinely considered only for polishing applications, such as was used at the previously described North Tahoe-Truckee Plant, where a low total or unoxidized nitrogen residual is mandatory. (Note that Tahoe-Truckee operates a granular activated carbon tertiary polishing system for organic nitrogen and carbon removal prior to its ion exchange/chlorination processes.) Use of sodium hypochlorite is preferred for safety reasons and because of the attendant buffer found with the caustic carrier (NaOH). Sodium hypochlorite adds about 1.7 mg of dissolved solids/mg of chlorine added.

One should anticipate subsequent residual chlorine removal when breakpoint chlorination is practiced. Activated carbon polishing for chlorine removal is complex and expensive and is not recommended over dechlorination by sulfite addition at larger facilities. Liquid sulfite addition (or possibly activated carbon) is recommended for small plants because of ease of handling and safety issues. Larger plants could use SO_2 cylinders; SO_2 hydrolyses in water to form HSO_3^-. The reaction of the sulfite with free and combined chlorine residuals is essentially instantaneous. Dosages are only 10 to 30 percent higher than stoichiometric predictions. The stoichiometric weight ratio of sulfur dioxide to chlorine (Cl_2) is 0.9:1. One mg of SO_2 will deplete 2.4 mg $CaCO_3$ alkalinity.

2.5.3 Non-Ion-Specific Technologies

The non-ion-specific processes are mentioned only in the interest of completeness. Here, all nitrogen forms are removed with other dissolved constituents with varying degrees of selectivity by the demineralization process of choice.

2.5.3.1 Reverse Osmosis

Reverse osmosis was selected as a representative demineralization technology. Six months of experience in the late seventies with reverse osmosis at Orange County's Water Factory No. 21, in California, suggested the following macrocontaminant removals (31):

Macro-contaminant	Influent, mg/L	% Removal	Residual, mg/L
Sulfate	338	99	3
Ammonium-N	22.5	94	1.4
COD	12.5	94	0.8
Sodium	158	89	17
TOC	6.0	82	1.1
Nitrate	2.9	55	1.3

The reverse osmosis system was preceded by the following processes: biological stabilization, high pH lime clarification, stripping (no aeration, just allowed to fall through the packing of the 24-ft tower), recarbonation and clarification, mixed media filtration, and activated carbon adsorption.

It is difficult to interpret the reported performance data for the reverse osmosis system other than to observe that nitrate appears more difficult to remove than ammonium and that the sum of these two nitrogenous compounds is near 3 mg/L after all of the preceding treatment processes. (Activated carbon would have achieved some removal of the soluble organic nitrogen.) Characterizations such as this lead to the conclusion that the only likely, and certainly the easiest, ways to achieve low-level ammonium or total nitrogen residuals are with biological processes. This is particularly the case if control of the macrobioavailable pollutants is the primary processing objective.

The 44-L/s (1 mgd) Denver, Colorado, secondary effluent to drinking-water quality demonstration pilot plant originally employed clinoptilolite to achieve its objective of less than 1 mg/L ammonium nitrogen. Clinoptilolite was reported to be unacceptably expensive ($0.40/1,000 gal) and management subsequently incorporated reverse osmosis for ammonium nitrogen control. Present findings suggest reverse osmosis ammonium-nitrogen removals on the order of 85 percent (32).

2.6 Key Fundamental Process Selection and Design Issues

Section 2.3.3 described the processes that are the principle focus of this manual. They embody, alone or in combination, suspended and attached growth treatment concepts in a higher technology, mechanical plant approach. This next section addresses key issues that influence the selection of the unit processes for the nitrogen control objectives.

2.6.1 Fundamentals and Empirical Factors

2.6.1.1 Comparative Differences and Needs

The purpose of this section is to provide a fundamental appreciation of the differences between suspended and attached growth technologies, and the ramifications of such in practice. Emphasis is placed on nitrification since it is the most frequently encountered design objective and without effective nitrification the best designed denitrification system has no chance for success. Finally, in terms of attached growth systems, emphasis is also placed on approaches found with trickling filter (TF) and rotating biological contactor (RBC) systems.

With respect to the system biomass, suspended and attached growth reactors are complete mix and plug flow systems, respectively. The suspended growth system's complete mix definition is derived from the return of settled biomass. Thus, save wastage, all of the organisms grown from treating the wastewaters are returned to contact the newly applied wastewater. Their relative activity at any point, however, will be controlled by the environmental conditions (e.g., DO), the applied substrate form and mass, and their history of exposure to the newly received substrate mass.

Conversely, in an attached film system without a return of the reseeding biomass or fluidization for mixing, the organisms found at any point in the vertical or horizontal processing train will reflect only the wastewater characteristics and environmental conditions that are routinely encountered at that point. Conceptually, it is difficult to visualize that the return of reseeding biomass to the fixed film reactor will impart significant benefit, since changes in the overall system solids residence time will be small. For any system that is required to remove appreciable amounts of $SBOD_5$, this small return may be accompanied by increased front-end DO stresses, with an accompanying risk of biomass accumulation and odor problems. However, if the system is called upon to nitrify, the recycle of the small amounts of nitrifiers encountered in the effluent from an attached growth reactor, or a fraction of the settled sludge from a downstream nitrifying activated sludge system, may improve nitrification performance. Some denitrification may also be achieved in a combined carbonaceous and nitrogenous oxidation system.

Aerobic metabolism in either attached or suspended growth systems is controlled by oxygen transfer to the liquid and, in turn, DO and substrate penetration into the bacterial film or floc. In suspended growth systems, the oxygen supply can be adjusted to satisfy the demand which, of course, is influenced not only by the applied substrate but also by the concentration of operating solids. The large volume (or long detention times) of suspended growth systems also serves to dampen the peaks and valleys of this demand, contributing to an inherent stability.

In a classical TF or RBC attached growth system, the oxygen supply potential is generally fixed by the static aerator nature of the media and transfer is only achieved by the magnitude of flow that passes over the media, or by the speed of rotation of the media through the fluid. The demand is experienced where it is experienced and is not subject to control by changing levels of operating solids concentrations. The limited volume of liquid (i.e., short detention time) encountered in an attached growth system contributes to its inherent conceptual instability.

The literature is mixed on the value of recycle to TF and RBC attached growth systems. Undoubtedly, that value depends on the system's application. With highly loaded systems, low SS recycle offers conceptual value in the transfer of more oxygen per unit mass of applied biodegradable substrate, less oxygen demand per unit volume of applied wastewater per unit time (giving load migration farther down the system), and additional scour of accumulated biomass (yielding more active surfaces). With low-loaded trickling filtration systems, low SS recycle offers some ability to mitigate such nuisance organisms as filter flies and snails as well as the potential of more effective exposure of the wastewater to all of the available surfaces through increased media wetting.

A flexible TF and RBC design approach would incorporate a capability for hydraulic recycle of low SS liquids to provide some ability for operating manipulation. An attractive operating feature for trickling filtration systems is the ability to control the localized hydraulic application rate independently of flow for nuisance control (i.e., excess biomass, organism, and odors) and improved load distribution (forcing it downstream). Similar approaches for RBCs would make use of an air scour for nuisance control and have the ability to cyclically reverse and step feed the flow through the process for improved load distribution and/or attached culture development.

Conceptually, the TF processing configuration is inherently inferior to the RBC because of its physical inability to distribute and cycle loads (except by serial application and change of the lead filter). A forced ventilation capability is highly desirable for a trickling filter system and is a recommended design approach.

Recent evolution of RBC technology shows that continuously operating supplemental air supplies are beneficial for heavily loaded systems. Submerged media, packed-bed attached growth systems directly use an air supply to respond to the system's oxygen transfer needs. Attached growth technologies with direct control of the air supply and certain usage of effective surfaces, not unlike suspended growth systems, yield the greatest certainty of predictable performance.

Once $SBOD_5$ loadings decline below those that yield readily observable nuisance conditions in classical TF and RBC systems, the controlling design considerations for carbonaceous stabilization are substrate transfer and oxygen transfer. Initially, in combined carbonaceous-nitrification systems, the dominance of heterotrophic flora and remaining carbonaceous substrate will control and use the available oxygen at levels that suppress any inroads by autotrophic nitrifiers. It should be remembered that nitrifier replication, in comparison to that of the organic stabilizing heterotrophs, is much more sensitive to DO concentrations.

The onset of nitrification is determined by heterotrophic/nitrifier competition and oxygen transfer considerations. For combined carbon and nitrogen oxidation trickling filters, experience suggests the following sequential activity through the system (33,34):

- Nitrifier growth is no longer limited by competition with heterotrophic organisms when soluble five-day carbonaceous biochemical oxygen demand ($CBOD_5$) values fall below about 20 mg/L.

- Once nitrification does begin, the nitrifying population builds to its maximum as determined by the oxygen transfer system, the declining significance of the stabilized wastewater's carbonaceous oxygen demand, soluble carbonaceous and nitrogen opportunity for diffusion into the biofilm, and the oxygen uptake of the combined aerobic heterotrophic and autotrophic biofilm population.

- When and where the maximum nitrifying population first occurs is difficult to identify beyond the certainty that it is probably accompanied by a soluble $CBOD_5$ on the order 10 mg/L.

- Thereafter, the maximum nitrifying population is controlled by the oxygen transfer characteristics of the reactor until the residual ammonium nitrogen drops below 5 mg/L.   *

- The nitrifying population then declines in proportion to the ammonium-nitrogen levels that are routinely experienced in the downstream reaches of the processing train and in relation to the ability to achieve sufficient biofilm for the population's retention.

Trickling filter systems will achieve consistently low ammonia objectives when properly designed and operated (35). The foregoing observations can explain why

classical attached growth system nitrification perform- ance shows little sensitivity to temperature over a range of 12–20°C (36,37) and why combined nitrogen-carbon oxidation and tertiary nitrifying attached growth systems have shown variability at low-level ammonium objectives and can have difficulty achieving consistent low-level am- monium residuals near 2 mg/L or lower.

The interrelationships of all factors point to the apparent complexity of fixed film process design. Seemingly unim- portant or inevitable plant practices, coupled with the nor- mal variability of incoming wastewater flows and pollutant concentrations, may have important consequences. Higher nitrogen concentrations (due to discontinuous sludge processing recycles and/or daily raw wastewater transient peaks) can result in unfavorable alkalinity to ammonium-nitrogen ratios or lower DO concentrations (associated with the competing spikes of carbonaceous material and the dominant heterotrophic population), which can suppress nitrifier activity at points where or- ganism concentration is highest. This forces the unoxi- dized nitrogen to points downstream in the processing train where the nitrifying population may be lower. Lower nitrifying populations, regardless of more favorable DO characteristics, will yield higher unoxidized nitrogen levels possibly beyond the target concentration. Poor develop- ment of biofilm and predation in the latter stages of the attached growth system, during average and low loading periods, further influences the ability of the system to perform under infrequent stress periods.

2.6.1.2 Coupled or Combined Processes

Blending attached growth trickling filter and suspended growth activated sludge technologies offers the designer the opportunity to take advantage of what each technol- ogy does best. Application of a suspended growth system with its returned solids after the trickling filter allows the designer to achieve superior flocculation, which, funda- mentally, is a function of the number of solids, time of contact, and mixing intensity. This allows trickling filter systems to compete with today's needs for rigid SS control. The trickling filter can also provide a roughing operation (or selector), with polishing provided by the downstream, now very stable, activated sludge system.

Recent or newly proposed advances in attached and sus- pended growth technologies continue this theme. The evaluator of coupled or combined new systems is best served by application of fundamentals, such as those described in the preceding subsection, and an awareness of past experiences with the operation of the individual processes. When compared to the limited understanding often encountered with emerging technologies, it is useful to remember that older systems and design approaches have been tempered by experience gained through years of practice.

2.6.1.3 Carbon-to-Nitrogen Ratios

Beyond the solids residence time concept, there is no more fundamental consideration with nitrogen control systems than what is termed "the carbon-to-nitrogen ra- tio." The purpose of this subsection is to introduce the reader to this important concept.

Figure 2-3 was prepared to provide a fundamental con- ceptualization of how the carbon to nitrogen ratio influ- ences the development of the nitrogen control strategy and its design response. Understanding begins with the characterization of the influent wastewater. As shown, raw wastewaters are relatively high in organic carbonaceous material and relatively low in unoxidized nitrogen; typi- cally, a COD:TKN of 10 to 15:1 is presented.

The biodegradability of the carbonaceous material in the raw wastewater is influenced by a host of variables, in- cluding the nature of the collection system, which also influences the inert SS in the influent wastewater. One certainty is that biodegradability will increase with dimin- ishing particle size. Those particles that escape primary sedimentation have the highest immediate biodegradabil- ity, while the retained particles must be hydrolyzed (solu- bilized) to simpler compounds before active cellular oxidation and synthesis can begin. Gray to black waste- waters signify that solubilization in the collection system is under way. Remote pumping stations promote active degradation by retention in the wet wells and particle breakup in the pumps. Combined sewers and major in- terceptors promote deposition and active decay of depos- its until they are resuspended under rain-swollen events. These are accelerated by warmer ambient and wastewa- ter temperatures.

Heavier, larger particles may or may not be separated prior to the biological reactor. If not, the carbon-to-nitro- gen ratio will be much higher than if they were. Once the substrate is applied to the biological reactor, the magni- tude of the resultant mixed liquor volatile suspended sol- ids (MLVSS) will be largely influenced by the applied refractory (or nonbiodegradable) particulate volatile mat- ter and the cell residence time of the heterotrophic micro- bial population.

The acclimated heterotrophs in the biological reactor will respond immediately to the soluble organics and yield both oxidation and new biomass end products. The MLVSS is composed of the applied refractory particulate organics, the slowly biodegradable particles derived from the applied wastewater, and newly formed cellular volatile matter. The slowly biodegradable particulate matter will also be eventually oxidized and synthesized into new cellular volatile matter. The new cellular matter will also be oxidized and resynthesized, leading to an accumula- tion of refractory cellular end products, the extent of which is dependent on the solids residence time of the system.

Applied
Substrate
(Raw Wastewater)

Organic Carbon	Nitrogen
Volatile Matter or COD[a]	[TKN[b]]

External
Substrate
Additive
Enhancement

High
Biodegradability

— Acclimation Needs?

Internal
Enhancement

Poorer
Degradability

Viable
Organic
Stabilizing
Heterotrophs

Denitrification
Issues

Settleable

Particulate

Colloidal

High   Soluble
Biodegradability

MCRT Dependent
VSS Synthesis, Respiration
and Resynthesis with
Refractory End Products

Net
VSS
Production

Nitrification
Limiting
MCRT

Net Initial Weight Gain
Due to Oxidation and
Synthesis of Soluble
Substrate

Organic Stabilizing
Heterotrophs

Nitrifying
Autotrophs

Typical Refractory or
Nonbiodegradable
Fraction (Not
Shown in Nitrogen
Distribution for
Clarity)

Organic N
Particulate

Soluble

Inorganic
(NH$_4$-N)

Net   Soluble Available and
Refractory Nitrogen

0   MCRT, Days

Nitrification Issues

Stabilized,
Oxidized
Substrate

Net
Carbonaceous
Substrate
Removal by
Oxidation

Remaining Viable
Nitrifying Autotrophic Biomass

Remaining Viable
Heterotrophic Biomass

Heterotrophic Refractory
End Products

Refractory
Material from
Original Substrate

Likely Net Nitrogen Removal Due to
Refractory and Synthesis
Removals

[a]BOD$_5$ is not presented because it is an
incomplete measure of the substrate and
its biodegradability.
[b]TKN (total Kjeldahl nitrogen) measures
the total unoxidized nitrogen.

No Remaining Soluble Substrate Biodegradability

Figure 2-3. Schematic of how carbon-to-nitrogen ratio influences development of a nitrogen control strategy along with the design response.

For all intents and purposes, appreciable concentrations of nitrifying autotrophs are not encountered until the satisfaction of some controlling solids residence time (θ_c) reflecting the concurrent environmental conditions that allow their replication and retention within the system. Once the limiting θ_c and environmental constraints are satisfied in a suspended growth reactor, the nitrifier population becomes dependent on the residual nitrogen that is available for nitrification (nitrogen available after correction for the amount removed by cellular synthesis, refractory particulate entrapment, and the soluble refractory organic nitrogen). In an attached growth system, as explained in Section 2.6.1.1, at any point in the reactor the nitrifier population will be influenced by competition between autotrophic and heterotrophic biomass.

This complex series of events is portrayed above the Nitrification Issues heading in Figure 2-3. As can be seen for the portrayed raw wastewater situation, the net viable autotrophic nitrifier population is small in comparison to the total net volatile suspended solids (VSS) production, as is the net nitrogen removed by synthesis. Clearly, to achieve a higher percentage of nitrifiers in the biomass, either the applied substrate carbon must be decreased or the applied available nitrogen must be increased.

Decreasing the applied substrate carbon can be readily accomplished, if desired (e.g., with primary clarifiers), and will lead to smaller reactors (for the same θ_c) and less overall oxygen demand. The applied C:N can also be manipulated in plug flow attached growth systems to affect a greater concentration of nitrifiers in the zone of ammonium concentration limiting performance (Section 2.6.1.1). Here, by cyclic switching of the lead trickling filter tower (if available), or by staging and flow reversals with an RBC system, higher overall nitrifier populations and lower ammonium nitrogen residuals may be promoted.

The requirements for achieving effective denitrification are the reverse of those encountered with nitrification. Again, the carbon-to-nitrogen ratio comes into play. Here, higher values of biodegradable carbon to nitrate nitrogen are sought to drive the denitrification reaction, and the viability of the heterotrophic population in the total biomass is a paramount concern for small reactor sizing.

52

The denitrification reaction can be positioned at the head, interstage, or end of a nitrification reactor or system. All the reaction requires is a source of nitrate nitrogen (which can be delivered by the return of a nitrified liquid), a heterotrophic biomass, and a carbonaceous substrate. The speed of the reaction and the extent of nitrate-nitrogen reduction are influenced by the magnitude and form of the carbonaceous substrate and its ratio to the oxidized nitrogen available for denitrification.

The complex series of events and processing alternatives accompanying the denitrification objective is schematically portrayed in Figure 2-3 over the Denitrification Issues heading. Unlike nitrification, the election of the denitrification process involves balancing a multitude of tradeoffs. The following paragraphs provide additional fundamental understanding.

The ideal substrate for denitrification is a readily biodegradable soluble carbon source, free of nitrogen, that is preferentially oxidized rather than synthesized. Methanol is such a substrate, but it is foreign to municipal wastewaters. As such, the organisms must be acclimated to it before it can be degraded. The acclimation response is dependent on the solids residence time and the methanol concentration. Culture isolation, whether it occurs in a suspended or attached growth reactor, yields the highest concentration of viable heterotrophs acclimated to methanol; but methanol leakage from the system still will occur if application rates suddenly exceed the mass to which the organisms have become accustomed.

When using foreign substrates to meet stringent, low-level nitrate-nitrogen goals, dosages must meet the highest applied nitrate-nitrogen mass and ensure that adequate substrate is present for all processing requirements. In order to avoid soluble substrate breakthrough, post-aeration of the mixed liquor suspended solids (MLSS) is used in suspended growth systems to compensate for overdosages, particularly when the available nitrate nitrogen is less than needed to satisfy the applied substrate. Soluble substrate breakthrough is more likely to occur in attached growth systems.

Practically, the most available substrate for denitrification at a wastewater treatment plant is the raw or settled raw wastewater. In this case, it should be remembered that all substrates are both oxidized and synthesized, and only that fraction oxidized will enter immediately into the denitrification reaction. The most suitable and immediately usable substrate in the denitrification reaction is the soluble biodegradable carbon. Thereafter, the denitrification rate proceeds as a function of the concentration and viability of the heterotrophs in the biomass and the biodegradability of the applied particulate carbon and the synthesized cellular products. Younger cultures promote a higher viability.

More slow to degrade, biodegradable particulate carbonaceous waste can be added through application of raw, as opposed to settled, wastewater. But this will lower the percentage of viable heterotrophic organisms in the total MLVSS and will increase the need for more oxygenation energy for carbonaceous stabilization. These two compromises may yield no net change in (or an even lower) reaction rate (e.g., mg NO_3-N reduced/mg MLVSS/d). Moreover, where high-level denitrification is required, it may be more cost effective to incrementally add a substrate such as methanol than to satisfy the additional carbonaceous oxygen demand associated with the raw wastewater.

Recent advances in combined nitrogen and phosphorus removal systems have sought to catalyze the biologically enhanced phosphorus removal process through the anaerobic formation of readily biodegradable volatile fatty acids. Volatile fatty acids formation, which is the first mechanistic step in anaerobic digestion, can naturally occur in collection systems, sometimes to a significant degree. Design approaches promote the formation of these acids through either a front-end anaerobic process or a sidestream fermenter.

With or without enhanced biological phosphorus removal strategies, the volatile acids generated from particulates also facilitate mainstream denitrification by increasing the C:N ratio. The sidestream fermenter approach, and subsequent liquid recycle, has particular conceptual attractiveness in that it minimizes the introduction of non- and slowly biodegradable organics (and inert SS) to the reactor and their consequent impact on reactor sizing. The negative aspects of the sidestream fermenter approach include additional unit processing commitments, a poorly stabilized sludge end product (needing further stabilization), lower potential for methane production, and the attendant nitrogen solubilization of anaerobic digestion, yielding the possibility of higher overall system oxygen demands.

The foregoing discussion reveals a multitude of complex considerations influenced by the carbon-to-nitrogen ratio. Fundamental understanding is only derived through the realization that it is both the magnitude and form of the carbon that count in nitrogen control systems. Differences in practice, as influenced by the collection system, and other unsteady environmental conditions and discontinuous plant practices account for the often significant variance observed in nitrogen control plant performance, even in seemingly similar situations.

2.6.1.4 Temperature

The impacts of temperature on process performance can be, and often are, of prime importance. Care should be taken when applying the results observed at one facility to another without understanding the temperature conditions. Care should also be exercised when applying tem-

perature adjustments in process kinetic characterizations, although it should be said that there is generally good confidence in process kinetic temperature descriptions of nitrifier growth rates.

The points made in the C:N ratio discussions should be recalled when addressing carbonaceous biomass predictions. The great bulk of the mixed liquor volatile matter is biologically inert and is not subject to change with changing temperatures. This explains why the reported net solids production remains largely constant with wide temperature changes and why for long-sludge-age systems it is not strongly sensitive to the solids residence time differences that are routinely encountered in practice.

2.6.2 Process Selection and Assessment of Design Issues

The material in the preceding subsection and the balance of this manual provides the basis for the qualitative comparative assessment of the suspended growth and attached growth technologies provided in Table 2-8. In using this table, the reader should bear in mind that it is only intended to provide general guidance. The often critical variable of site-specific considerations, as well as greater understanding and evolution of nitrogen control technologies with time, may make some of the furnished direction invalid.

Overall, the relative observations that can be made from Table 2-8 are summarized as follows:

- Suspended growth technologies are:
  - better understood,
  - less influenced by other processing considerations at the plant,
  - inherently more flexible and adaptable to a variety of processing objectives,
  - possibly a safer fundamental choice when high to stringent effluent standards are encountered.
- Attached growth technologies are:
  - often simpler to operate and maintain,
  - probably best applied in polishing or roughing roles with suspended growth technologies, or when the effluent objectives are not overly stringent.

2.7 Frequently Encountered Linked Design Issues

The beginning to this chapter introduces the concept that the appropriate unit process selection (or design) does not stand alone. It is influenced by all things that precede and follow in the overall scheme of the treatment works. The purpose of this section is to provide a fundamental and practical discussion of one certain and one likely design issue both of which are linked to the selection,

and often the sizing, of the nitrogen control process. These are solids-liquid separation and phosphorus removal, respectively.

2.7.1 Solids Liquid Separation

2.7.1.1 Suspended Growth Systems

A limiting consideration for sizing or operating a suspended growth reactor is its final sedimentation system. Often upsets occur because of one or more of the following:

- poor settling characteristics of the applied MLSS,
- too high design or operating MLSS loadings to the clarifier for the experienced flow regime, and/or
- inadequate or too low a return sludge rate for the applied MLSS concentration and its settling characteristics, leading to expansion and loss of the sludge blanket.

MLSS that demonstrate poor settling characteristics are commonly encountered in air activated sludge systems on a seasonal, if not endemic, basis. In municipal treatment, the major cause is the presence of filamentous forms of bacteria, which can flourish in the biomass because of low to trace DO levels in the mixed liquor or the sludge floc itself, or other factors. The older (or higher θ_c cultures, common in single-sludge continuous flow nitrogen control systems, inherently promote the retention of these filamentous forms. Other systems, such as the oxidation ditch, by the very nature of the isolated point of air (oxygen) application create reactor conditions that promote natural denitrification and the growth of filamentous forms. Similar situations may be encountered with mechanical aeration systems.

Although mitigated to a degree by raw sewage applications (because of its heavier organic and inorganic particles), settling problems abound in the historic application of both plug flow and complete mix variants of the activated sludge technology. Historically in these systems, the only available operating strategy was to reduce the operating MLSS concentration while trying to identify the cause of the problem. Reducing the MLSS to comply with the effluent SS objective surely compromises the reactor performance for denitrification, and may reduce the aerobic solids residence time to the point that seasonal ammonium or total nitrogen limits are violated.

Over the last 20 years, pure oxygen systems have been developed that can often reduce the variability and enhance the settleability and densification characteristics of MLSS. In this timeframe, and more specifically in the last decade, the technical literature has seen progressive interest and success in applying front-end selectors (aerobic, anoxic, and anaerobic) for improved MLSS settling and densification characteristics. For nitrification applications, it is worth considering pre-anoxic zones for their sludge volume index (SVI) and stability benefits. Pure

Table 2-8. Comparative Assessment of Suspended and Attached Growth Technologies Against Key Process Selection and Design Issues

Issue	Technologies*	
	Suspended Growth	**Attached Growth**
COMPATIBILITY: PREREQUISITE PROCESSES, EFFLUENT STANDARDS, AND SOLIDS PROCESSING TRAIN		
• Prerequisite Processes	None beyond preliminary treatment	Needs prior gross solids removal when applied to raw wastewaters
• Nitrogen Oxidation, Effluent NH_4-N		
— Intermediate, 5–10 mg/L	Easily achievable, difficult to control (consider split treatment)	Easily achievable
— High, 2.5–5.0 mg/L	Easily achievable	Achievable with good design
— Stringent, <2.5 mg/L	Good choice	May be poor choice without enhancements
— Variability	Lower	Likely higher
• Nitrogen Removal, Effluent Total N		All need external substrate addition; most installations in polishing applications.
— Intermediate, 10–15 mg/L	Easily achievable w/o external substrate	Easily achievable in most climates
— High, 5–10 mg/L	Achievable (normally w/o external substrate additions)	Likely achievable in most climates
— Stringent, <5 mg/L	Achievable (may require substrate additions)	Only fluidized bed and deep downflow bed demonstrated in cold climates
— Variability	Mixed, because of the upstream need for successful nitrification, often reliance on natural denitrification, and final variable of external substrate addition	
• Effluent SS Considerations		
— Clarifier for capture	Mandatory	Depends on application and technology
— Effluent SS consistently <10–15 mg/L	Filter becomes increasingly mandatory	Depends on application and technology
• Vulnerability to Intermittent Sludge Processing	Lower because of longer hydraulic detention time	Higher because of shorter hydraulic detention time (consider equalization and bleedback, especially with anaerobic digestion)
INHERENT PROCESS FLEXIBILITY		
• Adaptability to Maximize Internal Denitrification	Excellent	Poor to nonexistent as presently demonstrated
• Adaptability to Variety of Operating Control Strategies	High	Low
• Adaptability to Biologically Enhanced Phosphorus Removal	High	Not directly
• Wastewater Temperature Influences	Nitrification and denitrification kinetics	Often not controlling for nitrification with some technology applications, applicable to denitrification kinetics
• Principal Controlling Process Variables	Reactor volume, operating solids concentration, oxygen transfer	Reactor volume (media-specific surface), hydraulic loading, and oxygen transfer

Table 2-8. Comparative Assessment of Suspended and Attached Growth Technologies (continued)

	Technologies*	
	Suspended Growth	**Attached Growth**
SENSITIVITY TO NEEDS OF OPERATION		
• Operating Simplicity	Inferior	Superior to same
• Ease of Maintenance	Inferior	Superior to worse
• Vulnerability to Upset as a Result of		
— High Flow	Higher to same	Lower
— High Raw Pollutant Mass	Lower	Higher
— High Raw Pollutant Concentrations	Lower	Much higher
— Deterioration of Recycles	Yes, if sustained	Much higher
— Toxic Compounds	Varies	Varies
• Recovery from Upset	Mixed	Mixed
• Ease of Process Troubleshooting	Better	Worse
• If Plant Is Noncompliant, Is the Fix:		
— Solvable by Operations?	At times	Less frequently
— Solvable by:		
Minor Capital Improvements?	Often	Not as frequently
Major Capital Improvements?	Not as common	Often
• Aesthetics		
— Sight	Generally superior	Generally inferior
— Sound	Potentially inferior	Generally superior
— Odors	Generally superior	Potentially inferior
— Nuisance Organisms	*Nocardia* scum or float	Flies, snails, and worms
ADDITIONAL THOUGHTS ON PROCESS SELECTION		
• Present Applied Understanding of Nitrogen Control Technology		
— Oxidation	Generally superior	Generally inferior
— Removal	Generally superior	Generally inferior, except for warm weather
• Plant Site	All sites	Often attractive with high natural gradients

* Avoid rigid thinking. Best process selection may be a blend of technologies, both coupled and uncoupled, using each for what it does best.

oxygen and selector technologies are largely directed to mitigation of the immediate soluble substrate impact on the MLSS, in terms of natural selection of filamentous forms at low DO concentrations. Process development and proprietary technology considerations with nitrogen control applications undoubtedly served as a catalyst for many packaged technologies offered by various suppliers. In some processes, such as the sequential batch reactor, it is relatively simple to establish the selector concept when choosing the operating sequence.

Today, the following guidance seems appropriate for control of poorly settling and densifying activated sludges:

• Beware of reliance on high MLSS designs (see next paragraph).

• Provide some form of selector technology.

• Still assume that uncontrollable bulking may occur and provide the ability to add chlorine or hydrogen peroxide to the return sludge, so that the operator has an immediately implementable external control procedure.

The design MLSS issue is, ultimately, a function of the applied solids concentration and mass loading to the clarifier, and in this respect there are no hard and fast rules, only general guidance. Ideally, the MLSS concentration should be set to encourage nonhindered, as opposed to

hindered, settling solids characteristics, a breakpoint that often occurs at about 2,000 mg/L for air activated sludges and at times some 50 percent higher for pure oxygen activated sludges or sludges encountered with selector technologies. The breakpoint is specific to the characteristics of each sludge. Generally, settling velocities with MLSS concentrations below these values show little gain, while settling velocities associated with MLSS concentrations above these values rapidly deteriorate.

Care should be exercised in assuming that acceptable solids loads are constant (since they decrease with increasing MLSS values once hindered solids settling conditions are reached) and that the clarifier solids flux models are complete (since they ignore effluent SS, their probable statistical variation and the sludge blankets that must be achieved to match the desired return sludge concentrations). Simple solids flux models (when applied to circular units with suction pickup) do work productively, as several field scale trials have proven. If scraper mechanisms are used, solids flux models break down as the effective thickening area is reduced. Without upstream flow equalization or diversion, or unless the design is for relatively low peak overflow rates, conventional air activated sludge system designs that anticipate MLSS concentrations more than 3,000 mg/L under the elevated flow regimes are at risk unless firm supportive data suggest otherwise; in no case should designs with MLSS concentrations in excess of 4,000 mg/L be judged acceptable unless the clarifier has been sized specifically to handle these high values. With anoxic-aerobic systems, good SVIs of 80 to 120 mL/g have been consistently demonstrated; this easily allows MLSS concentrations of 3,000 to 3,500 mg/L with reasonable clarifier designs. Similar criteria for pure oxygen (and possibly selector technologies) are advocated by some to be 30 to 50 percent higher, but this is not generally agreed upon.

The return sludge rate and assumed return solids concentration are probably two of the most generally abused design and operating parameters found in today's practice. Rates must be set to meet needs and do not simply reflect an arbitrary percentage of flow. Conceptual identification of needs starts with the understanding that, in the ideal situation, there is no sludge blanket. With air activated sludges, best operation (lowest effluent SS) is typically achieved with a return sludge solids concentration of around 5,000 mg/L. Without flow equalization, designs that rely on a return sludge concentration in excess of 7,500 mg/L for a sustained period of time should be used with caution. Designs that need a return sludge concentration in excess of 10,000 mg/L for more than a day should be judged generally unacceptable. Again, similar criteria for pure oxygen (and possibly selector technologies) are sometimes 30 to 50 percent higher, but even these systems can have problems with poor settling sludge.

The guidance in the preceding paragraphs will serve the user well in general practice. Unacceptable excursion of elevated MLSS concentrations will be avoided by the provision and use of a reliable, tolerant sludge handling system and disposal outlet. This guidance inherently assumes that the purpose of the clarifier is to clarify not thicken. Sidestream thickening, exclusive of storage of waste activated sludge, yields the greatest assurance of mainstream processing success.

2.7.1.2 Attached Growth Systems

Attached growth systems also have their solids-liquid separation problems. These problems are found with difficult to settle colloidal suspensions and sudden, sustained sloughing from the reactor. The following paragraphs address these issues.

Colloidal suspension problems are best addressed by remembering the principles of flocculation. Effective flocculation is sensitive to several variables: solids contact opportunity, time, nature of the material to be flocculated, and mixing energy. The opportunity for solids contact is most important and, with trickling filter and rotating biological contactors, can be achieved through a flocculator that is external or internal to the clarifier, with a settled solids return. The design range of operating solids in these solids contact systems should be more than 500 mg/L and no more than 1,500 mg/L if there is no other reason to maintain a higher operating level. Contact times should probably be no less than five minutes at the controlling maximum flow regime. The solids retention time is also important. One day is typical for TF (plastic media) systems; longer aerated solids contact detention times or the use of reaeration tanks may be required. Because the floc is weak, a flocculator center well is often used to compensate for breakup in the transfer line. Full-scale studies have shown improvements in supernatant SS of 5–15 mg/L by use of the flocculator center well (38). Energy values should be on the order of 50–100 m/s-m (fps/ft).

The uncontrolled sloughing problem that is frequently encountered with trickling filters (more so than with rotating biological contactors) can be related back to organic loading conditions and inadequate flushing rates. High organic loads promote the formation of dense growths of organisms in the lead media sections of the treatment unit. These growths can build up to either block the media openings or tear from the media and completely or fractionally end up in the system's effluent. Excessive sloughing can quickly overwhelm a downstream reactor; problems at the clarifier are less certain. If the downstream reactor is called upon to nitrify, performance may be compromised as a result of oxygen transfer limitations or lower nitrifying solids residence times induced by the influx of solids.

Avoiding sudden sloughing of the biomass is best done by promotion of a more continuous, controlled sloughing. One approach is to increase the recycle rate. As described in Section 2.6.1.1, if this is an inadequate or unavailable option, a flow distribution system can be considered that is independent of processed flow for TFs and RBCs. With an RBC system, an air scour capability can be incorporated.

2.7.2 Phosphorus Removal

The following subsections highlight considerations that may be encountered when applying phosphorus removal in concert with nitrogen control technology. The reader desiring more information is referred to the MOP 8 (1), MOP FD-7 (39), and the Design Manual for Phosphorus Removal (25).

2.7.2.1 Metal Salt Addition

Metal salt addition for phosphorus removal is commonly encountered throughout the United States at facilities with and without nitrogen control. Its impacts on the design of nitrogen control systems are the result of additional solids production and alkalinity demand. Aluminum salts make less sludge than iron salts, but often do not dewater as well; the alkalinity demand for both is the same on a molar basis. Table 2-3 provides stoichiometric information relating to these two processes.

The designer and user are reminded that phosphorus immobilization (whether by metal salts or biological enhancement) causes the plant's SS phosphorus content to increase (often double) and that excessive dosages, beyond 1.5 to 2.0 moles of metal ion per mole of available phosphorus (less background and synthesis removals), do little to enhance soluble phase phosphorus removal if there is no accompanying decline in the pH. Often, plants, especially the small ones, add more metal salt than necessary for reliable phosphorus removal.

In alkalinity-poor wastewater, metal salts are best added to the effluent from the nitrification reactor. Here, target alkalinities of 20–30 mg/L as $CaCO_3$ are acceptable, and, if approached, may yield phosphorus removal benefits due to the declining pH. Otherwise, the target residual alkalinity from the nitrifying reactor should be at least 50 mg/L as $CaCO_3$.

The addition of iron salts results in more precipitated material than aluminum salts. The chemical feed system should be sized to deliver both. The ideal point of addition is to the biological reactor effluent before clarification. Split treatment, (i.e., using multiple dosing points located at unit processes designed for solids-liquid separation) often yields lower overall use and higher performance. Such strategies may be · appropriate with suspended growth systems (applications to the primary clarifier influent and MLSS effluent) to minimize the inert solids that are carried in the biological reactor.

The appropriateness of metal salt applications to a primary clarifier that is followed by an attached growth reactor is conceptually questionable, since the formed inert solids that carry over from the primary clarifier may mask active media surface areas. However, such concerns may have no practical significance given the abundance of more significant variables that influence process performance.

Polishing dosages of metal salt (alum more often than ferric) can also be added ahead of any final effluent filter to achieve both capture of colloidal fines that would otherwise escape the filter and additional phosphorus removal benefits. Metal salts and polymer provide an excellent means of achieving a low SS effluent.

With suspended growth systems, combined or separate metal salt and polymer application strategies with primary treatment are always worthy of consideration when cold-weather nitrification requirements can result in large downstream reactors because of the high biomass that must be carried to meet the longer solids residence time requirements. This approach is especially attractive if design flows are uncertain and are substantially in excess of present flow conditions, or if an existing plant's non-compliance problems are restricted to colder months.

2.7.2.2 Biologically Enhanced Systems

Biologically enhanced phosphorus removal has advanced through the last decade to the point that in all cases it is "worthy of consideration for design," which means that the technology must stand or fall on its own merits. Ultimately, it may be substrate limited (both for forcing the reaction and for waste solids production) and nearly, if not completely, dependent on how the waste solids are handled in the solids processing train.

Biologically enhanced phosphorus removal needs the formation of fatty acids. These can be best generated in an advance anaerobic contactor or a sidestream anaerobic fermentor. Fatty acid generation also can occur in the denitrifying anoxic reactor (paralleling anaerobic reactions within the interior of the sludge floc or biofilm), but only with greater, albeit acceptable, conceptual risk. The wisdom of an anaerobic fermenter in terms of nitrogen removal considerations is discussed in Section 2.6.1.3.

Biologically enhanced phosphorus removal technologies fit naturally with suspended growth single-sludge systems designed for carbonaceous removal; but they are also compatible with nitrifying systems. Their concepts are also used with anaerobic and anoxic selectors to avoid bulking sludges. The process's adaptability to attached growth systems is just emerging, but would appear to necessitate some period of SS contacting, and likely use of an off-stream acid fermenter.

Low-level influent phosphorus concentrations may make either biological enhancement or metal salt addition the

technology of choice. A high-level phosphorus influent may make metal salt additions mandatory whether or not enhanced biological phosphorus removal strategies are incorporated.

With biological enhancement, the incremental phosphorus waste solids production numbers are some 55 and 75 percent of the values predicted for iron and aluminum metal salts, respectively. These waste solids are inherently more unstable than the precipitated metal hydroxyphosphate solids, since the additional phosphorus removal is a stress-induced, temporary cellular storage product. Some of the enhanced biologically removed phosphorus returns to solution under extended periods of aeration (mainstream and digestion), and short (hydrolysis) and long periods (stabilization) of anaerobic treatment. Ultimately, the success of the technology is clearly a function of how the biologically enriched, phosphorus-laden solids are processed and removed from the system.

The ideal, generalized processing train would incorporate direct wastage to a dissolved air flotation thickener, followed immediately by solids dewatering and stabilization of the concentrated product. All other approaches, in some way or another, present the opportunity for additional compromise. Sometimes the beneficial formation of magnesium ammonium phosphate will occur during anaerobic digestion. This, however, may replace the solubilization problem with a scaling precipitate, a problem that appears to be most commonly associated with extremely long digestion detention times and may correlate with the mineral hardness of the native water.

The present level of knowledge about this technology suggests that, although solubilization does occur, it is rare that the phosphorus removal efficiency would be reduced to levels equivalent to background synthesis values (about 0.02 mg P/mg VSS). Given the present level of uncertainty found in practice, a metal salt addition capability would be recommended for backup. The safest design of the biological reactor for nitrogen control would downrate, if not ignore, the phosphorus removal credit assigned to biologically enhanced phosphorus removal. One would then assume that the required degree of phosphorus removal might still have to be achieved by metal salt addition and account for the resultant inert SS in the reactor. Alternatively, if there is competition for the available wastewater carbon between biological phosphorus removal and biological nitrogen removal, direct methanol addition should be considered to satisfy the carbon demand for denitrification. This would minimize (or eliminate) the need for chemical addition (and its associated sludges) in conjunction with phosphorus control.

2.8 Nitrogen Control Troubleshooting and Problem-Solving

Throughout this chapter, an attempt has been made to illustrate the concept that "everything is connected to everything else." In nitrogen control system troubleshooting and problem-solving, the investigator is equally served by remembering that "all things ultimately have to be what they are, not what they are assumed to be." In the real world (40):

• Each treatment plant is absolutely unique.

The variables are almost limitless: e.g., customers (yearly, seasonally, weekly, daily, hourly variability), collection system (flat, steep, expansive, small), climate (ground water, precipitation, temperature), plant processes (mainstream and sidestream, hydraulic and solids detention times, overload, underload), auxiliaries, pumps and blowers (use and nonuse, down for repair, out of service), sound and well-intended operating strategies (continuous and discontinuous, attended and nonattended), recycles and solids disposal.

• Process models often are regression (empirical) models of no mechanistic import and no general validity.

Empirical is not a dirty word. It is the translation of the experience of practice, or observation of the end result, to describe or predict complex things. They are site specific and are as good as the extent and quality of data used to develop them.

• Mechanistic models are valuable teaching tools but usually cannot be verified and the cause-and-effect relationships underlying the observed phenomena may not be known.

Be aware of their guidance but beware of the "apparent" validation. Remember that each treatment plant is unique; model constants are derived from empirical observations and refined through the implicit regression of experience.

If a real nitrogen control problem is encountered, reread Tables 2-2 and 2-4 and Sections 2.6.1 and 2.7 (and Sections 2.9.3.3 and 2.9.3.4 may also be helpful). Then, depending on the problem, read Chapters 3 (nitrification) and/or 4 (denitrification). The applicable experience reported in Chapters 6 through 8 may also be helpful. Then consider reading Chapter 5. While being mindful of the introduction to this section, nitrogen control or related troubleshooting and problem-solving can begin.

Beginning with some general guidance for problem-solving, the following subsections describe some of the most commonly encountered problems with nitrogen control technology. All but one are associated with nitrogen oxidation, since satisfaction of this objective is key to nitrogen removal by denitrification, and the majority of the design applications exclude nitrogen elimination. Often the problems can be anticipated and mitigated in design.

2.8.1 General Guidance

Troubleshooting and problem-solving start with the acquisition of knowledge. This knowledge is used to understand and define the problem and its cause (or causes). Solutions readily flow once the problem and the cause are identified.

Regardless of the particular situation, problem-solving follows from a readily identifiable sequence of activities. Table 2-9 provides step-by-step guidance for wastewater treatment plant problem-solving. Its applicability is broad. The five offered steps are always necessary. Thereafter, the questions and comments may not always be appropriate because they depend on the problem at hand. The questions and comments under each step are the ones that have been generally found to be appropriate. They are presented with a view to stimulate. It is recognized that completeness is an unobtainable goal.

2.8.2 Microbial Inhibition and Pollutant Breakthrough

2.8.2.1 Toxicity

Microbial toxicity due to inhibitory compounds is probably the most discussed but least real problem encountered in municipal wastewater treatment. Section 3.3.5 of this manual provides specific information on a limited number of inhibitory compounds and their effects on nitrification kinetics. Reference 1 provides additional information. Industrial pretreatment programs can, and are legally required to, address this problem when one exists. The community that consciously fails to do so is as legally liable as the industry.

This is not to say that microbial inhibition does not occur, but only to point out that the cause more often than not is environmental (e.g., pH and DO), not inhibitory wastes. If gross microbial toxicity is suspected, a quick check of the operating DO level and BOD_5:COD of the raw wastewater and that applied to the reactor may determine if it is a valid postulate. Under acute toxic conditions, the reactor's operating DO level could suddenly go up (doubling or more) and the BOD_5:COD could suddenly go down (50 percent or more) from background levels. If both are not observed, acute microbial toxicity is unlikely. Microscopic examination of the mixed liquor could also be helpful.

The BOD_5:COD check may not apply for toxic upset of nitrifying organisms, which are generally recognized as the most sensitive of the flora routinely encountered in wastewater treatment. However, toxic microbial inhibition could still be demonstrated by a sudden rise in the reactor's DO with an accompanying significant increase and decline in the plant's normally occurring ammonium and nitrate-nitrogen levels, respectively. This sensitivity of nitrifying organisms to toxics, in toxic amounts, holds conceptual promise as a continuous bioassay or biomonitoring device. Chronic toxicity, such as experienced by

heavy metal accumulation, will develop slowly and may not be fully realized prior to 2 to 3 times the solids residence time.

2.8.2.2 Soluble BOD5 Breakthrough

Two causes that yield a soluble BOD_5 breakthrough are inadequate oxygen supply and inadequate biological treatment for the received biodegradable substrate.

The oxygen supply is rarely the cause of soluble BOD_5 breakthrough, especially in suspended growth systems. Oxygen transfer efficiencies increase with declining DO values. The biomass has some sorptive affinity for soluble substrates and heterotrophic organisms retain a soluble substrate removal capability in the absence of measurable DO levels, through either anoxic (e.g., denitrification) or anaerobic pathways. This leaves inadequate biological treatment as the most likely culprit for soluble BOD_5 breakthrough.

Inadequate biological treatment for the received biodegradable substrate starts with the solids residence time of the reactor and ends with the substrate and its activity and sources in the collection system. The reactor's response to any unique foreign substrate, like methanol (see Section 2.6.1.3), is only as good as the time and dosage (mass) provided for acclimation. A discontinuous release of a unique foreign biodegradable substrate (e.g., as with an industrial processing run or an end-of-the-week cleanup) can mean the absence of an acclimated culture that may have responded in some fashion to the last discontinuous release.

2.8.2.3 Ammonium Breakthrough

Ammonium breakthrough from a plant that appears to be safely designed on the basis of solids residence time (or rate) is generally associated with one or more of the following: inadequate oxygen transfer, inadequate alkalinity, too high an ammonia concentration for the acclimated biomass, solids washout, and toxicity.

Inadequate oxygen transfer is more a problem with attached rather than suspended growth reactors, but can be experienced in both. For problem-solving, it is well to remember that nitrification cannot occur without DO and that nitrifier activity can be highly variable when measured operating DO levels decline below 2 mg/L. The gross DO measured in the fluid is only an indirect measure of what is actually available in the interior of the biological floc or film. Organic load spikes or diurnal swings act to suppress and delay nitrifier activity when the DO is depleted by the more dominant response of the viable heterotrophs (see Sections 2.6.1.1, 2.6.1.3, and 3.3.3). Fundamentally, the solids residence time of the system is of no consequence to the nitrifiers if a sufficient period of proper oxygenation is not available to ensure their replication.

Inadequate alkalinity ultimately leads to a pH depression. Nitrifiers can and do acclimate to lower pH conditions, but

60

Table 2-9. Step-by-Step Recommendations for Wastewater Treatment Plant Problem-Solving

WHAT IS THE BASIS FOR SAYING THERE IS A PROBLEM?

Avoid Stupid Mistakes—Check analytical procedures and quality assurance/quality control.

Analytical kit procedures often differ from more rigorous analyses. Are checks by an independent laboratory appropriate? Does an independent laboratory know what it is doing? Does the problem coincide with a change in analyst or analysis? Does the problem lie within the variability of the test? Is there a need for triplicates? Do operations set up or conduct tests on weekends, while the laboratory people do this on weekdays? Are BODs performed with inhibition of nitrification? Are BODs reseeded after chlorination?

Is the Overall Problem Understated or Overstated?—Check sampling.

Often a problem is understated. Weekends are often the worst time [Friday (end of the week) industrial cleanups, sampling and operation most relaxed; worst impact from solids processing train, since what enters the plant on Friday, leaves the plant processes on Saturday (or Sunday)]. Twenty-four-hour composites mask the severity of breakthrough and dumps. Are samples manually or automatically collected? Single grabs or composite grabs? What is the compositing period? When are samples collected? What are the procedures? Are reported samples and flows what they are identified to be (reported raw flows often include recycles, reported raw wastewater often includes recycles)? Remember nitrogen transformations (organic nitrogen hydrolizes to ammonium; nitrite-nitrogen is encountered in both oxidative and reducing systems). Do automatically collected sample containers overflow? Are final effluent samples collected after chlorination? Guard against preferential or lax sampling and failure to report weekend conditions.

WHEN WAS THE PROBLEM FIRST OBSERVED, AND WHAT WAS DONE ABOUT IT?

Often the problem was not observed when it first occurred, or the reasons for the problem have something to do with something that happened in advance of the problem. Look for subclinical signs, several months before the problem was first reported. Remember the great buffer found in the liquid returns and recycles, hydraulic volumes, and solids (the process and plant sludge age) found at the plant. Remember that mixtures take about 2.2 residence times to come to 90-percent equilibrium with new stimulants or operations.

Determine remedial cures that have been tried and the results of these tries. Believe the data more than the personal interpretations. Understand what the words mean (e.g., turned on, speeded up, faster, tried). Remember people present themselves in the most favorable light and sometimes, unknowingly, suppress valuable information until the right question is asked.

DO PLANT OPERATIONS CONTRIBUTE TO THE PROBLEM?

Look particularly at all discontinuous operations and the reality of all recycles as they are experienced in the main flow stream. Check solids mass out the door (effluent and to disposal), against solids mass in (raw wastewater and inert solids formed from added chemicals). Don't overreact to sidestream pollutant concentrations until their mass significance is known. Prepare approximate mass balance for guidance. Perform inert solids balance. Understand what the processes see (pollutant mass and concentration) as they actually see it, and how the process responds to it (e.g., complete mix or plug flow). Know what equipment was down and why. Know if artificial constraints were imposed (e.g., air supply, pumping, chemical feeds). Determine factors about operation that make equipment operate the way it does.

IDENTIFYING THE CAUSE OF THE PROBLEM: RESPECT EVERYONE'S OBSERVATIONS, MISTRUST EVERYONE'S INTERPRETATIONS

Remember that the right solution already may have been tried, just not long enough, and that people are smarter and less smart than they appear. Often the correct solution to the problem has been identified by someone; often the person doesn't know it. Talk to everyone, ask what they observe, and see what they think. Avoid preconceived ideas. Be open to all. Try to develop a scenario that matches all observations. Do not eliminate any observation that doesn't fit the scenario until you can confidently do so. With elimination of irrelevant observations and retention of all others, the remaining scenario probably correctly identifies the cause(s) of the problem. Expect that it is related to more than one thing, and catalyzed (or most observable) by only one to a few.

Mistrust yourself. Remember the uncertainties and check concepts by the total wealth of experience. Say when you don't (and the field doesn't) know. Get help if still confused, or if it is logical to expect that somebody knows more about a critical something than you do. It is illogical to expect anyone to know everything.

SOLUTIONS TO THE PROBLEM: EASY TO IDENTIFY, OFTEN HARD TO IMPLEMENT

Until demonstrated for a year or more, all solutions have risk, some more than others. Try to develop solutions that don't have the potential of making anything worse. Remember that bigger is not necessarily better. Seek a rifle, not a shotgun. Consider everything that can be done in the collection system (operationally), then everything that can be done structurally. Prioritize in terms of certain benefit. Define risks of each in some qualitative way. Think of ways to minimize risk. Develop fallback procedures before they are needed. Think of costs as if your money were at stake.

Share this entire thought process with decision-makers. Identify key issues and concerns. Determine what results constitute a fully acceptable solution. Identify options to achieve desired results. Agree on a remedial program implementation strategy. Take remedial action steps with a clearly defined flexible plan of action, with known fallback positions and points of possible irretrievable misuse or loss of resources that can serve as new decision points.

not excessively so (usually less than 6.0 to 6.5). When low pH conditions are encountered, the nitrifiers are inhibited in the rate of their reaction. If inhibited too frequently, the total nitrifier population will decline because of washout, regardless of the operating DO. In diagnosing pH problems, use grab samples before flows drop over weirs (this fall can cause the pH to rise, as can the routine laboratory procedure of shaking the sample thoroughly before performing any analysis, due to CO_2 stripping; see Section 2.5.1 for more understanding). Places to look for alkalinity deficient, low pH problems are discontinuous recycles and acid dumps in the collection system. Routine monitoring of the effluent alkalinity often provides valuable insight as to irregular occurrences.

Closed carbonaceous oxidation systems, such as encountered with pure oxygen technology, can contribute to low pH problems since CO_2 stripping cannot occur. Similarly, fine-pore aeration systems may have a lower pH than coarse bubble systems, and rotating biological contactors may have a lower pH than trickling filters. Ultimately, the pH of any system with adequate alkalinity for nitrification will reflect its equilibrium with the CO_2 produced and the atmosphere, as determined by the physical opportunities for CO_2 stripping.

2.8.2.4 Oxidized Nitrogen Breakthrough

The anoxic denitrification reaction is accomplished through the use of driving substrates (externally added substances or untreated wastewaters) and/or endogenous respiration of cellular material. Oxidized nitrogen breakthrough is generally associated with one or more of the following: particularly low applications of the driving substrate, failure to acclimate to the dosage of the driving external substrate, competition for the driving substrate, lower rates than expected, and physical limits of the process. Responses are self-evident and essentially translate to provide more substrate, more time for acclimation, spread the substrate out to more places of application, and use more solids and/or reactors (see Section 2.6.1.3). Tracking of the soluble COD, organism oxygen uptake rates, and oxidized nitrogen forms into, through, and out of the reactors will provide valuable insight for problem-solving.

2.8.3 Suspended Growth Systems

2.8.3.1 Unwanted Seasonal Nitrification (and Bulking)

The elimination of nitrification can be just as important as its promotion for cost-effective treatment. Strategies for its elimination are the reverse of its promotion. They begin with the direction to drop the solids residence time to a point that is acceptable to satisfy the applicable processing objectives but is insufficient to sustain nitrification. Waste solids production levels may marginally increase, but it is nearly always cheaper to process and dispose of the extra sludge than to use oxidation energy for its destruction.

Methods for dropping the solids residence time include dropping the MLSS. Do not be afraid of lower values than those found in the textbooks; but for most sludges, go no lower than 500 mg/L because of flocculation concerns. For sludges with high SVIs, even values below 500 mg/L can work well with a goal of maintaining an initial settling velocity no higher than about 3.6 m/hr (12 ft/hr). Other strategies involve taking reactors out of service, provided oxygen transfer capacity is adequate in the remaining reactors. The best design is the one that allows this to be done without loss of the downstream clarification capacity.

Do not over aerate and be careful if the solids residence time drops below three days since bulking sludges might result. Consider the use of return sludge chlorination for specific nitrifier kill. At low MLSS concentrations, remain sensitive to the possibility of detergent foam and provide (or use) the spray water capability.

Most municipal wastewater treatment plants probably operate at a θ_c beyond their needs; especially those that are designed for secondary treatment equivalency and are significantly below their design load. Often, inroads of nitrifiers occur in the spring, acting to depress the DO, and result in seasonal bulking. If this happens, remember why. Do not unnecessarily add more air; instead, reduce the MLSS and solids residence time. Find a safe reasonable minimum for the plant, its service area, and the effluent objectives.

2.8.3.2 Scum (Floatables) Management

Progressive use of high θ_c, nitrifying single-culture activated sludges over the last two decades has also brought the coincidental observation of *Nocardia* froth. Perhaps coincidentally, this time period has also seen the mandatory regulatory requirement of secondary clarifier baffling and scum capture.

The designer should anticipate *Nocardia* froth, or process-generated floatables, with combined carbonaceous oxidation/nitrification activated sludge systems. It is not generally observed with low-sludge-age air systems, nor isolated nitrification or denitrification suspended growth systems. It may be endemic with pure oxygen systems and their submerged reactor effluent drawoffs. Excessive froth formation is often encountered in the combined sludge anoxic reactors with overzealous use of baffling. Scum (or floatable) problems with attached growth systems or coupled applications of an attached growth system followed by suspended growth reactors have not been reported to date.

Process floatables are best managed by the philosophy that seeks to avoid their return to the processing train by separate processing through the point of ultimate disposal. If this is not possible, then they should only be slowly introduced, after concentration, to the sludges delivered to the dewatering process. In the liquid processing

train, the obvious key is to avoid froth entrapment until planned for. The first likely place for entrapment is at the mixed liquor effluent channel. Here, consider using its natural flotation to assist in its removal. Downstream, the sedimentation tank should have full-width tilting troughs. Elsewhere, at all other upstream and downstream points of entrapment, consider covers so scum can be seen or allowed to escape, and the use of sprays where it might be observed. A number of facilities report some success in chlorinating the spray water.

2.8.3.3 Fine-Pore Diffuser Clogging

The rising popularity of fine-pore diffusers may bring an accompanying problem with clogging. Cost analyses generally always show that very frequent cleaning can be justified (41).

The clogging problem appears to be most severe where the applied load is the highest and the DO levels the lowest. The problem may be exaggerated with combined carbonaceous and nitrogenous oxidation systems because of the collective mixed liquor oxygen demand of the combined heterotrophic and autotrophic bacteria. Clogging further exacerbates the problem at the front end of the reactor and the inadequate DO levels transfer it to the later stages of the system.

Beyond cleaning, the following mitigation measures are possible: drop the operating solids level, increase the air supply, use selector technology, use coarser diffusers or alternate oxygen dissolution equipment where the oxygen demand is greater, or spread the load across the length of the aeration tank. Designs should anticipate the possible need for frequent cleaning of the front-end diffusers in plug flow reactors and always provide for convenient access and rapid drainage of aeration tanks.

2.8.4 Attached Growth Systems

2.8.4.1 Oxygen Transfer

Understanding attached growth systems starts with understanding oxygen transfer potential and experienced oxygen demand characteristics (see Section 2.6.1). The oxygen supply is largely fixed by the reactor and its media. The dissolution is determined by the fluid applied to it, the fall, the driving deficit, and, in the case of RBCs, by the speed of revolution of the contactor. The uptake is influenced by the localized concentration of biodegradable substrate (both nitrogen and carbon) and localized concentration of heterotrophs and nitrifying autotrophs. The microbial population at any point in the train will reflect more the average conditions than discontinuous peaks, but will respond to any transient condition.

Heterotrophic reactions, if possible, will dominate nitrifying reactions because of less sensitivity to DO, faster growth rates, and their greater number. Temperature impacts on nitrifiers will only be experienced after the achievement of nonlimiting DO conditions. Heterotrophic activity will be catalyzed at a faster rate in response to temperature impacts since heterotrophs can perform aerobic, anoxic, and anaerobic reactions as required and will be made possible by the depth of growth and the biodegradability of the material within the growth.

The oxygen demand imposed by the film will vary according to current and historical conditions. Throughout the year, the entire system will be in a state of flux responding to the localized concentration of pollutants and other environmental conditions such as temperature, DO, and pH. Often the lowest seasonal flows will coincide with the warmest temperatures, yielding concurrence for the most limiting reactor conditions (highest localized reactor oxygen demands per unit time) with the highest nitrogen concentration, at a time when effluent standards are the most stringent.

Nitrifier activity will follow wherever the heterotrophic activity will allow. The exertion of the nitrogenous oxygen demand will be controlled by the number of nitrifiers encountered in the processing stages. The population will be limited first by concurrent heterotrophic oxidation, then by the oxygen transfer of the system, and then by the available ammonium nitrogen.

Although heterotrophic biomass is not essential for nitrifier attachment, the heterotrophs form biogrowth to which the nitrifiers adhere. Biogrowth formation and attachment will become progressively more fragile and spotty as the available carbonaceous substrate and heterotrophs decline. Biogrowth formation in the earlier stages of a combined carbon oxidation/nitrification process will build, and will either fall off and/or block void space used for oxygen transfer. This has the potential for permanently or temporarily blocking flow and air passages, creating new random flow patterns, and can serve to create another complex set of conditions influencing the performance of the entire system.

The overall "simple to operate and maintain" attached growth system belies the complexity of the reactions that occur within it. Precise attempts to predict performance may be futile. Remedial strategies for the noncompliant system will surely encompass one or more of the following:

- equalization of loads (upstream or as applied to the reactor),

- split treatment to increase substrate to starved heterotrophs in order to form more biogrowth that the nitrifiers can adhere to,

- instream and/or isolated sidestream cycling of reactors to force more ammonium to where the biomass sees little (consider controlled digester supernatant as a source of ammonium to force more growth of nitrifiers; be careful of alkalinity needs),

- cycling load reversals where possible to maximize nitrifier growth,

- higher recycle flows for dilution of the soluble substrate and higher oxygen transfer, and/or

- forced ventilation.

2.8.4.2 Biogrowth Control

Technology exists to control excessive biogrowth in combined carbon oxidation/nitrification systems. For a trickling filter, the ideal situation may be the ability to control the localized application rate independent of the flow and recycle through an electric drive on the distributor arm. If not available, the next best approach may be to slow down the arm by reversal of the distribution ports. Conceptually, this approach may bring maintenance needs in conflict with the retention time needed for performance. For a rotating biological contactor, maintenance needs are easily satisfied with air scour.

Both approaches reflect a common desire to scour off excess biogrowth before it becomes a problem, thus achieving a constant controlled sloughing, rather than living with the unpredictable whims of the system and the unpredictable impacts on the biological culture and the physical media.

Insufficient biogrowth can only be resolved by the introduction of additional substrate. Controlled additions should assist in the formation of suitable conditions for retention of the autotrophic nitrifiers. Strategies include higher localized flow rates, split treatment, and isolation and submergence.

2.8.4.3 Nuisance Organisms

A high degree of nuisance organism control is achieved by the scouring procedures described in the previous subsection. A capability to chlorinate the recycle water also seems appropriate.

Some systems, particularly nitrifying systems, seem to be bothered by snails. The degree of control that can be achieved by scouring procedures or media selection is uncertain. Peoria, Illinois, has had some success in controlling snails in its nitrifying RBCs by bypassing a portion of the primary effluent directly to the affected RBC units. The best strategy is to assume that snails will occur, and to provide a baffled spot with a depression and sump in the line immediately after the attached growth system for their capture and accumulation. The simplest management scheme for the captured snails may be to deliver them to the head of the plant's degritting operation.

2.8.5 Transient Chlorination Demands

A common problem encountered at many wastewater treatment plants is a sudden loss of the chlorine residual and a dramatic increase in the chlorine required to achieve either a regulatory set or an operationally set residual. This demand is associated with both chlorine oxidation of nitrite nitrogen and breakpoint chlorination with low ammonia levels.

The nitrite-nitrogen transient chlorine demand is typically encountered at plants that chlorinate year-round and move from non-nitrifying to nitrifying conditions. It is also encountered in nitrifying plant startup. During these conditions, the nitrite nitrogen is present only briefly (as much as a week or two) as an intermediate point (or temporary operating condition) in establishing a stable nitrifying population. Appropriate operating strategies range from living with it (i.e., placing the chlorination system on manual and not worrying about the chlorine residual as long as the required level of disinfection is met), to getting out of the condition as fast as possible (i.e., going to a younger solids residence time operation to eliminate nitrification, or dropping back on the sludge wastage as much as possible to promote the rapid creation of a stable nitrifying population).

Nitrite-nitrogen transient chlorine demand problems promise to increase in frequency as more polishing denitrification technologies are applied to wastewater treatment plants, since nitrite-nitrogen is an intermediate step in the denitrification reaction. Its presence here is symptomatic of too little carbonaceous material to drive the denitrification reaction to completion, or inadequate acclimation or reactor contact time. In this application, sudden excursions of the chlorine demand can be used as an immediate alarm that something may be wrong at the denitrification step.

At disinfection doses of about 10 mg/L of chlorine, a plant averaging 1 mg/L of ammonia nitrogen will go into breakpoint chlorination during significant portions of the day and experience periods with no residual chlorine and no disinfection. Some plants have had to overdose chlorine (30 mg/L) to overcome the problem which is more common than the nitrite problem.

2.9 The Design Examples

2.9.1 Introduction

Sound process design concepts and considerations are best illustrated through specific examples. This section was prepared with this objective in mind. Its purpose is to illustrate the design activities that precede the detailed evaluation of any unit process and to provide a common design condition for the unit process nitrogen control design examples in Chapters 5 through 8.

2.9.2 Treatment Facilities for the Design Examples

The great majority of U.S. wastewater treatment plants now (and anticipated in the future) range from 4 to 440 L/s (0.1 to 10 mgd), with about 65 percent of the plants treating about 30 percent of the flows. Figure 2-4 schematically characterizes "simple" and "more complex" ge-

Plant A (Simple)   Plant B (Complex)

Flow Equalization   Flow Equalization

Preliminary   Preliminary

                                                         Primary

Biological   Biological
Reactor   Reactor

Secondary   Secondary
Clarifier   Clarifier

                                                         Filtration

Disinfection   Disinfection
$(Cl_2/de\ Cl_2)$ Thickening   Thickening $(Cl_2/de\ Cl_2)$

                                           Anaerobic
                                           Stabilization

Storage   Second-Stage
                                           Storage

Dewatering   Dewatering

Curing/   Curing/
Storage   Storage

                        Land
                        Application

Figure 2-4. Basic schematics of "simple" and "complex" wastewater treatment facilities with a typical flow range (4 to 440 L/s).

neric wastewater treatment facilities that may be typically encountered in this flow range. Other than the possible provision for alkaline stabilization of the dewatered sludges to address newly promulgated sludge disposal criteria, the processing schemes are believed to illustrate the technologies that are routinely incorporated into plants of this size for levels of performance equivalent to secondary. These facilities (designated as Plants A and B, respectively) provide the basis for design examples found in this manual. Detailed process flowsheets are presented in Figures 2-5 and 2-6 for the "simple" and "more complex" process trains, respectively.

2.9.2.1 Commonalities

Inspection of the process flowsheets provided in Figures 2-5 and 2-6 shows a common approach for pretreatment and minor residuals management (screenings, grit, and floatable removal and processing), the possibility of flow equalization, disinfection and detoxification (dechlorina-

tion), ample solids storage (solids equalization), waste activated sludge thickening, and dewatering and ultimate disposal with the firm backup of landfill disposal. The principles demonstrated in these schematics, with respect to these considerations, are as follows:

*   Although small in average quantity (typically, no more than 3–6 g dry solids/m^3 [25–50 lb/Mgal] raw wastewater), minor residuals are highly variable, large in significance, and often a troublesome and demanding consideration for design and operation. After their removal, the designer and the operator are challenged as to what to do with them. The designer should plan for their subsequent handling and processing with the same care as the major residuals found at the treatment facility. As shown, adding lime will prevent nuisance conditions and provide control of infectious agents. If at all possible, design and operation should attempt to blend these materials into the ultimate disposal plan for the major residuals generated at the

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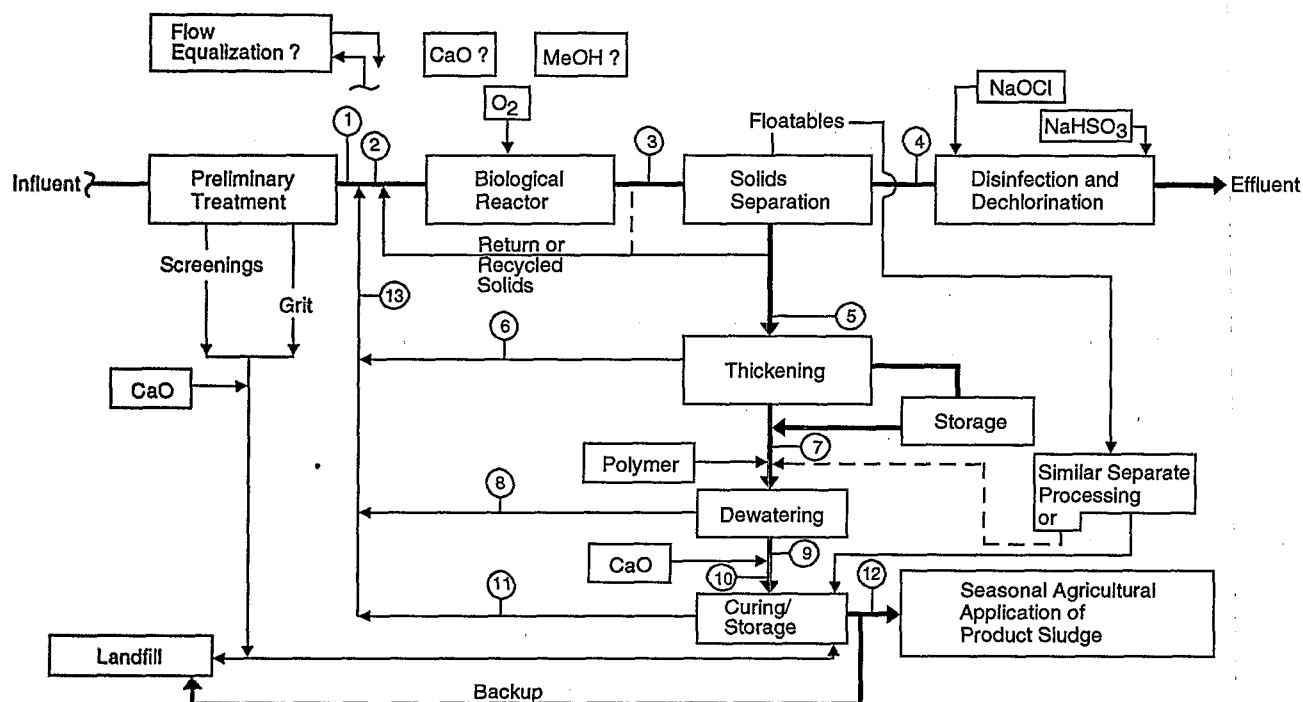
Note: Circled numbers indicate process points for mass balance calculations. Note also that schematic is not applicable to attached growth systems, because they lack all forms of primary treatment.

Figure 2-5. Detailed schematic of "simple" wastewater treatment facility (Plant A) (see Table 2-15 for mass balance data).

plant site. This is a possibility for the grit. If this is not possible, then there is no other recourse than to landfill; often onsite burial is used.

- Particular care should be taken in managing floatables derived from the processing train. Process-generated scum is often particularly troublesome with suspended growth systems. Raw scum can be readily concentrated on standing, or by screens, to concentrations in excess of 20 percent solids. Process scum also has a natural flotation to yield solids similar to what would be expected from a dissolved air flotation thickener (and this is an excellent point of reintroduction to the sludge train if the thickened sludge is immediately dewatered). In anticipation of the accumulation of troublesome floating residuals, suspended growth liquid and solids process reactor design should avoid submerged points of withdrawal. Raw wastewater floatables and those generated in the biological processes (e.g., *Nocardia* scum) are best managed by separate processing up to the point of ultimate disposal. This avoids their reintroduction into the processing train. If this is not possible,. they should only be slowly introduced, after thickening, to the sludges delivered to the dewatering process. When ultimate disposal of the waste solids for beneficial use is anticipated, both the waste sludges and the floatables should be ground to eliminate the

visual observation of unsightly rubber and plastic articles common to municipal wastewaters.

- With so much of the liquid processing train design ultimately dependent on hydraulic considerations, flow equalization is an attractive concept for process stability, especially with smaller plants and those that contemplate suspended growth treatment technology. The small plant designer and owner should seriously evaluate this consideration.

- The choice of sodium hypochlorite and sodium bisulfite for disinfection and dechlorination reflects the desire for safety (easily handled liquids) and minimal risk in what may be unattended operations (or a poorly trained and/or supervised staff) at small plants. Additionally, where nitrification is required to protect the oxygen resources (and biota) of the receiving stream, the attendant need for dechlorination is likely. In practice, use of Cl_2 and SO_2 is also commonly observed. Ultraviolet (UV) radiation disinfection has become a broadly accepted alternative to chlorination, particularly when dechlorination would be required. It is well demonstrated, does not require handling of chemicals, and leaves no residual. The designer should seriously consider UV in lieu of chlorination/dechlorination in advanced plants (nitrification or greater) and/or plants that incorporate filtration.

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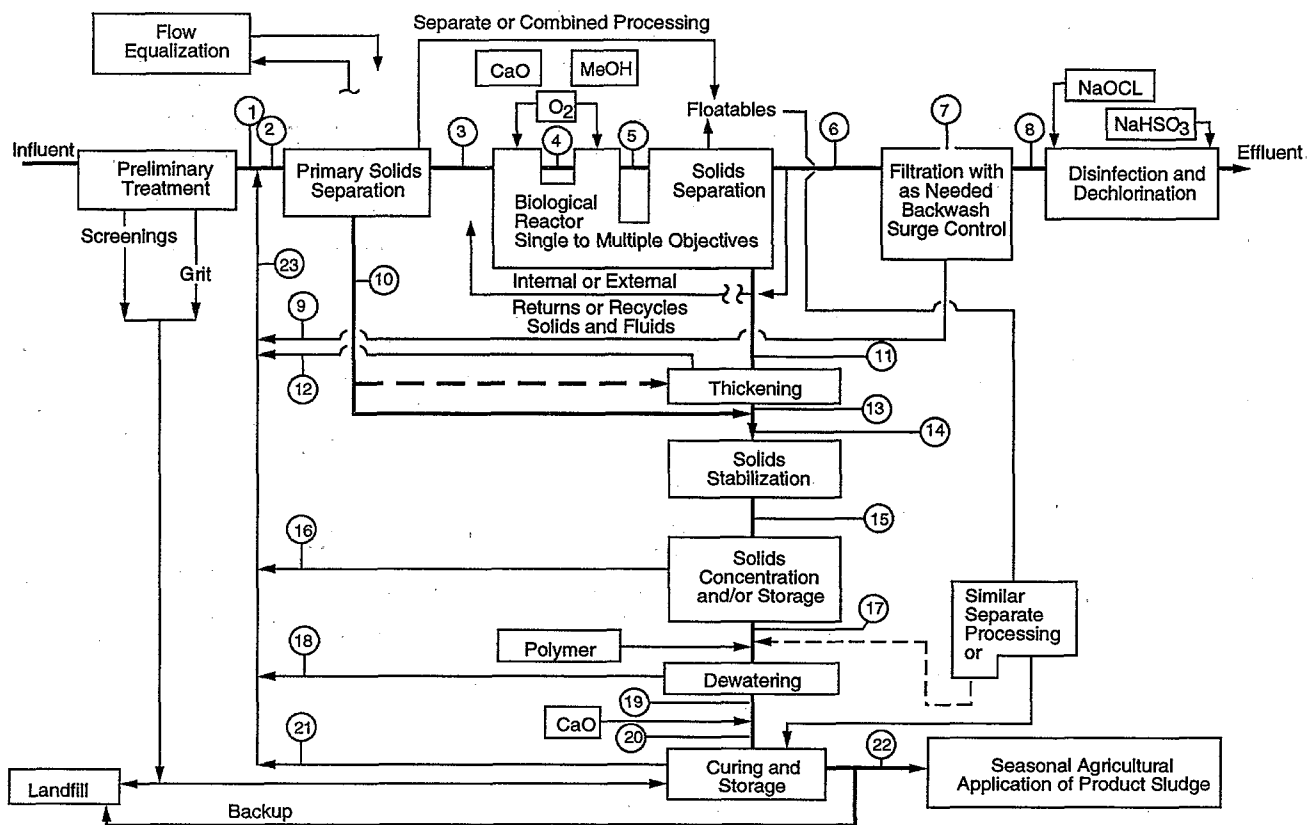
Note: Circled numbers indicate process points for mass balance calculations.

Figure 2-6. Detailed schematic of "complex" wastewater treatment facility (Plant B) (see Table 2-16 for mass balance data).

- Often a point of plant failure results from attempting to achieve too high a return sludge concentration and the resultant elevated sludge blanket found in the clarifier. Both the simple and more complex plants show the use of a separate thickening process for the waste activated sludge to highlight the importance of side-stream concentration of this process stream. Attempting to concentrate the waste activated sludge in the primary clarifier can often lead to washout of these lighter solids into the secondary system during higher flows. This consideration is particularly important with nitrogen control facilities where sudden influxes of solids from the primaries may cause a washout of the nitrifiers due to effluent losses or the need for excessive wastage. If concentration in the primaries is attempted, lower surface overflow rates than routinely allowed by regulatory guidelines may be appropriate. With multiple primary installations, alternate strategies may include flow peak control for select primaries that receive the waste activated sludge.

- Solids storage systems provide the same service as flow equalization in the liquid processing train as well

as providing strategic responsiveness to uncertainty. In terms of equalization, the solids storage system must be matched with the planned operation of downstream operations. In small plants, the frequency of dewatering operations may range from monthly or seasonal intervals to a fraction of an eight-hour shift per week. Where agricultural applications are anticipated, solids applications are limited to preparation for the one or more crops that are harvested per year. Lime stabilization of the dewatered solids product and wind-row curing for a year or more solids production may be reasonable with some beneficial recycle objectives.

- The designer should pay particular attention to odors if lime stabilization is selected. Ammonium will dissociate to ammonia at elevated pH conditions. This is particularly important when processing raw primary or anaerobically digested waste solids.

- It is important to note the backup of landfill disposal in the processing schematic. Sound plant design and operation must provide the assurance that residuals derived from wastewater treatment can be eliminated.

67

This is best done by not assuming that the preferred way of ultimate disposal will be fully realized.

2.9.2.2 Differences Between the Two Plant Scenarios

Practically speaking, the principal difference between the two plants is only the presence of the primary solids separation step (gravity sedimentation) in Plant B. The introduction of primary sedimentation allows a more optimized approach in terms of tankage commitment (and lower possible construction costs) but only for a more demanding operation (and higher possible operating costs). This, of course, is the crux of wastewater treatment plant design: unit process optimization versus operating convenience and cost minimization.

- Primary sedimentation is often linked with an isolated anaerobic solids stabilization (digestion) system. The solids stabilization process shown in Figure 2-6 is assumed to be anaerobic digestion in order to illustrate the often troublesome impact of digester recycles. Solids stabilization in the simpler Plant A occurs in the mainstream reactor by aerobic processes, reflecting a solids residence time which would yield a well-stabilized sludge (although still restricted in beneficial use scenarios).

- The reader should note that the Plant A schematic specifically excludes attached growth system applications because of the absence of primary clarification, which is necessary to avoid fouling and possible clogging of the media. In the more complex Plant B, the required level of protection is classically achieved by gravity sedimentation, although some form of finer screening than found in the preliminary treatment phase of the process may also be worthy of consideration. Fine screening may present a more demanding operation and, at the time of this manual's preparation, is rarely encountered in U.S. municipal wastewater treatment applications.

- The mainstream biological reactor in Plant B can accommodate more varied configurations than the simpler Plant A, including either a suspended growth or fixed film process. It can contain one or more sludge cultures or mixed applications of suspended and attached growth technologies, in a coupled or uncoupled manner. Filtration is also shown in the Plant B process schematic, addressing final effluent polishing. Consideration of the filter backwash is worthy of further discussion.

- When backwashing is discontinuous, it can represent a significant bump in the plant's flow; the processing schematic in Figure 2-6 illustrates the use of a surge tank for equalization, as needed. The question of where to return the filter backwash is a design decision. The processing schematic shows the filter backwash returned to the head of the plant for convenience, but, practically, it could just as easily be returned to the upstream secondary clarifier (this would avoid the hydraulic surge through all of the plant but lose the peak mitigation which is inherent in the reactors and conduits of the liquid processing train). However, the filter backwash is also a source of seed organisms from the upstream processes. In the case of a nitrifying system, it may be appropriate to return it to the lead carbonaceous oxidation reactor, and provide an inoculum of nitrifying bacteria throughout the processing train. Such practices, using filter backwash and/or waste sludges, are commonly encountered at multi-culture treatment plants.

- Finally, detailed inspection of the Plant B schematic shows internal processing stream callouts for the biological reactor (Callout 4) and the filter (Callout 7). The purpose of Callout 4 is to allow a fuller understanding of the internal workings of the biological reactor(s) as they may be applied to meet either internal or overall processing objectives. The purpose of Callout 7 is to demonstrate the process response to methanol additions for denitrification within this unit process.

2.9.3 The First Design Steps

Process design follows a series of logical steps. The first design steps are described in the following paragraphs. The introductory chapters of the recently revised MOP 8 provide further information and detail for the interested reader (1).

2.9.3.1 Understand Processing Objectives

The first step in any process design is to understand the processing objectives. These objectives, from an overall plant perspective, include the:

- effluent limitations,

- needs (or standards) for residual solids beneficial use or disposal, and

- various considerations that influence the design, including:
  - future planning and service,
  - plant and adjacent area aesthetics (sight, sound, and smell), and
  - O&M expectations and realities.

Collectively, the understandings associated with these processing objectives feed backward into the design of the integrated facilities and the specific unit processes found at the treatment facility.

The processing objectives of the design examples are provided in Table 2-10. Two different effluents are presented. In both cases, process calculations are to be preformed for controlling wastewater temperatures of 10°C (50°F), 15°C (59°F), and 20°C (68°F). The first set of effluent limits establish typical secondary effluent criteria for BOD_5 and TSS, whereby the maximum 30-day

Table 2-10. Effluent Objectives of the Design Examples

	Effluent Limits (<Avg. Concentration, mg/L)	
	Maximum Month, 30-day	Maximum Week, 7-day
Effluent 1 (seasonal)		
CBOD$_5$	30	45
TSS	30	45
NH$_4^+$-N	2	3
Total N	10	15
Effluent 2 (year-round)		
CBOD$_5$	10	15
TSS	10	15
NH$_4^+$-N	2	3
Total N	5	7.5
Total P*	1	1.5

*Present design implications of phosphorus limit.

and 7-day average concentrations must not exceed 30 and 45 mg/L, respectively. Nitrogen control is required on a seasonal basis (generally May through October) and limits are set that can be met through relatively easily achievable control strategies. The maximum 30-day average total nitrogen and ammonium nitrogen limits are 10 and 2 mg/L, respectively.

The second set of effluent limits is more stringent and imposes advanced wastewater treatment requirements with respect to BOD$_5$ and total suspended solids (TSS). A total nitrogen limit of 5 mg/L is required in this case on a year-round basis. The design examples will also present the impact of a phosphorus limit (1.0 mg/L) on the nitrogen control process design.

General considerations for reliable solids disposal are distributed throughout this chapter. The other considerations that influence plant design are beyond the scope of this manual unless they are specifically associated with some aspect of the nitrogen control technologies. These considerations are discussed in this chapter and the process-specific chapters that follow.

2.9.3.2 Understand Influent Wastewater Characteristics

The important influent wastewater characterizations in nitrogen control include quantification of the flows, pollutant loads (mass and concentration), and other parameters pertinent to the process design at all conditions of interest. The conditions of interest always include the compliance interval associated with the plant's effluent pollutant limitations and should be explored for each season if seasonal restrictions apply, or if institutional, industrial, or

such special wastes as septage loads can be expected to vary with the season.

The effluent compliance intervals defined in the plant's permit represent the regulatory agency's implicitly desired statistical reliability for the treatment facility. For example, the maximum month and the maximum week per year correspond to a statistical recurrence interval of about 92 and 98 percent of the time, respectively. The design of the treatment facility (and the unit processes therein) will be controlled by the maximum week or month in the most restrictive season. The controlling restrictive season may change per the pollutant under consideration. For example, the design of the nitrogen oxidation system may be dictated by the late winter/early spring wastewater temperatures, whereas the maximum organic and solids load may occur in the summer because of seasonal tourist or industrial increases in the plant's service area.

Experience has shown that the average day effluent pollutant concentrations (corresponding to a median or 50 percent occurrence value) will typically be about one-half to one-third of the more extreme conditions encountered at the maximum month and week. With no information to the contrary, the designer should elect processes which, either individually or collectively, can routinely achieve residual pollutant concentrations far superior to the conditions defined in its discharge permit.

The conditions of interest to a successful process design are not limited to various maximum value characterizations. Minimum conditions are also important for the design of turndown capabilities and early operating year strategies. Minimum conditions are also important when assessing the performance of processes which rely on linked biochemical reactions for successful control of various pollutants. For example, when the denitrification process relies on influent wastewater as its carbon source and must meet stringent total nitrogen effluent restrictions, the design may also require competing carbonaceous substrate demands associated with enhanced biological phosphorus removal. Pollutant concentrations at minimum flow should also be checked for their ramifications in process design (particularly important with attached growth systems) and compliance with concentration based effluent standards.

Clearly, the sound design of the integrated works addresses both maxima and minima conditions and takes the necessary steps to assure that reliable performance will be achieved at each. Ideally, site-specific raw wastewater maximum and minimum information is desired, but there is only limited information in the readily available literature (42,22).

Common sense guidance, under dry weather conditions, suggests that maximum soluble nitrogen concentrations will precede flow peaks and that ratios of the maximum nitrogen mass loads to average nitrogen mass loads are

likely to be some 10 to 25 percent greater than the ratios of the dry weather flow peaks to average day conditions; similarly, minimum nitrogen mass loads are likely to be some 90 to 75 percent of the corresponding dry weather flow minima. The relationship between ammonia peaking and hydraulic peaking loads for several cities is presented in Figure 2-7. Seasonal infiltration and event-specific inflow would be expected to have little influence on the mass of nitrogen received at the plant (except that the latter may bring in previously deposited suspended matter in various degrees of stabilization from the collection system).

Tables 2-11 and 2-12 summarize the influent wastewater characterizations elected for the design examples. When developing such characterizations do not allow precision to overwhelm common sense. Nothing in municipal wastewater treatment is ever known with greater validity than within 10 percent of the correct answer, and variability is the norm.

The average day characterization is presented in Table 2-11. The values presented are believed to be representative of a typical municipal wastewater in the United States without any unique industrial, commercial, or institutional contribution. The serviced population is estimated at about 6,250 individuals per 44 L/s (1 mgd) of average contribution. The average per capita flow reported in EPA's 1986 Needs Survey was about 0.6 m³/person-day (160 gpcd) (43). Average annual daily per capita SS and BOD_5 emissions in the example are about 0.09 kg (0.20 lb) and 0.10 kg (0.23 lb), respectively. Note that U.S. domestic per capita releases are on the order of 0.26 m³/person-day (70 gpcd); smaller communities tend to be less subject to infiltration and inflow and, if free from significant industrial contributions, may reveal significantly lower overall per capita flow rates and slightly lower per capita pollutant emissions.

The volatility of the influent SS reflects the assumption that the influent BOD_5 is lower than the SS, and the assumed infiltration condition derived from ground and storm waters in the collection system. Higher influent SS volatilities would be expected with tighter collection systems and a strongly domestic service. Higher wastewater alkalinities are often encountered when the community's raw water supply is from ground-water sources. References 1 and 39 provide additional background material for wastewater characteristics and nutrient contributions, respectively.
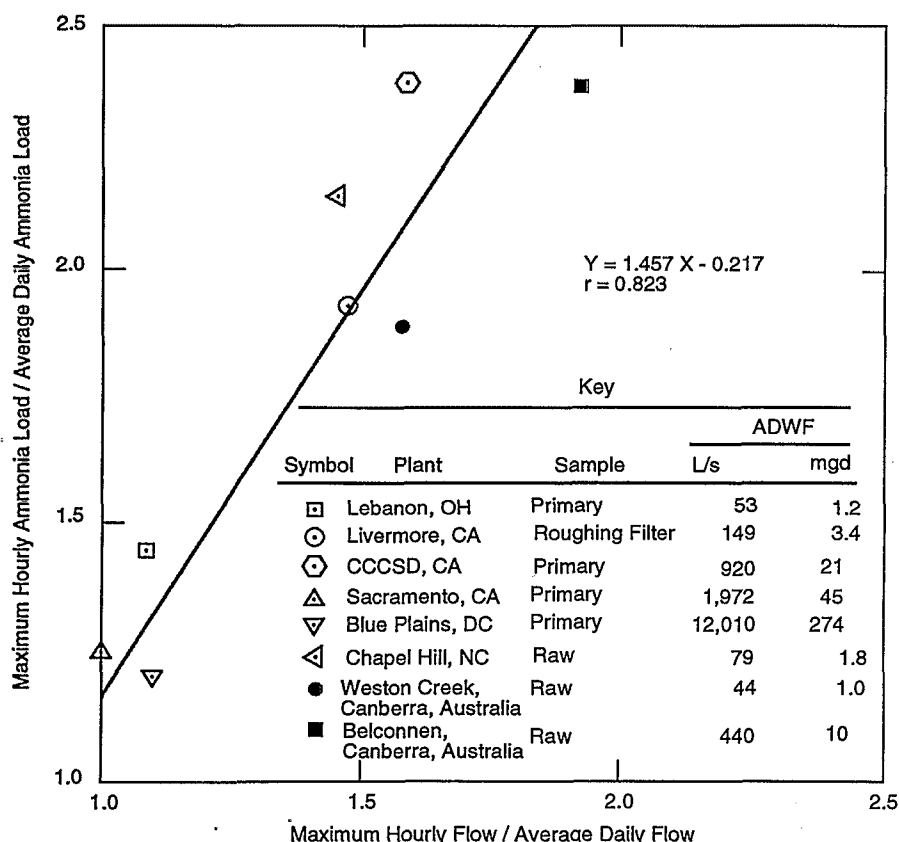
Figure 2-7. Relationship between ammonia and hydraulic peaking loads for treatment plants with no in-process equalization (from Reference 22).

Table 2-11. Design Examples: Average Day Raw Influent Wastewater Characterization

Parameter, mg/L	Inert	Volatile*	Total	Soluble	Particulate*	Total
SS	52	123	175			
%	30	70	100			
$CBOD_5$				50	100	150
COD				100	180	280
Total Nitrogen				24	6	30
Ammonium Nitrogen				20	—	20
Organic Nitrogen				4	6	10
Total PO_4-P				5	1	6
Alkalinity, as $CaCO_3$				120	—	120

* With the exception of the BOD_5, 30 percent of the volatile SS and the related particulate pollutants (COD, TKN, and PO_4-P) are assumed as nonbiodegradable.

Table 2-12. Design Examples: Influent Wastewater Peaking Factors

Condition	Percent of Time Conditions*	Flow	SS and Organics	Total P and N	Matching Alkalinity
			Ratio of Noted Condition to Average Day Pollutant Mass		
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			

* Equivalent percent of time conditions are less than or equal to stated values.

Table 2-12 provides a reasonable characterization of the peaking factors needed to complete the process design. For the purposes of design, it is assumed that the maximum and minimum conditions can happen at any time during the year, and that minimum and maximum pollutant masses may or may not coincide with minimum and maximum flow regimes. Proper design should identify and evaluate the controlling condition (e.g., maximum load with minimum or maximum flows).

2.9.3.3 Understand Performance Characteristics of Existing Operations

Knowledge of the processing objectives and the influent wastewater characteristics converge at the operations of the existing or planned plant. With an existing plant, the designer should make sure that full understanding of all operations is achieved, including such fundamentals as:

- the sampling and monitoring program used to describe the plant (including the presence or absence of any recycle at any point of sampling or measurement),

- the actual and desired duration of intermittent operation such as sludge wasting, processing, and disposal,

- the performance and quality characteristics of the solids handling processes in terms of their feeds, product solids, and, if possible, product liquids, and

- validation of any perceived bottlenecks and limitations which may be described by operations and/or the experience of the field.

The above information should be used to describe the performance of the plant under existing conditions (and those in the future) and help identify compliance strategies and likely attendant improvements (including those which offer more beneficial or optimal use of the existing works) to meet the processing needs of the future while, if not overcoming the problems of today, at least not contributing to their unacceptable exacerbation.

2.9.3.4 Define Process Constants and Variables

The preceding analyses allow the designer to address areas of uncertainty and adequacy. Decisions are needed with regard to the following:

- the adequacy of the existing works and appurtenant equipment under the present and future operating strategies,

- the attractiveness of alternate operating strategies to realize improved process performance, flexibility, and/or additional processing capacity,

- whether or not alternate unit processes for the same processing objective need be evaluated, and

- the important design constants and variables for each process found in the possible integrated processing trains.

The decisions that are reached at this point guide the preparation of the mass balances described in the next subsection. As can be inferred, the nitrogen control system decision is only one of many, and not necessarily the most important, to be reached by the design engineer.

One of the more critical variables needing definition is the biological reactor solids production. This characterization begins with an understanding of the carbonaceous removals that can be anticipated under normal operating conditions. Given freedom from significant dumps of unique biodegradable or nonbiodegradable industrial wastes, as associated with discontinuous industrial operations and end-of-the-week cleanups, Figure 2-8 provides a reasonable characterization of the likely soluble $CBOD_5$ and soluble COD removals as a function of the mean solids retention time, θ_c (or mean cell residence time, MCRT) in the biological reactor. Each plant is unique as represented by the ranges shown in Figure 2-8. Under non-oxygen limiting conditions, Tampa achieves over 95-percent $SBOD_5$ removal at θ_c near one day. Jenkins and Garrison (44) note that it is very difficult to detect variation in soluble degradable effluent COD for plants operating below substate removal rates of 3 kg COD removed/kg VSS/d. The likely θ_c for designs which provide for secondary treatment equivalency (the maximum month SS and $CBOD_5$ objectives of 30 mg/L, without nitrification as a design objective) and for nitrification are shown for reference purposes.

As shown in Figure 2-8, soluble COD removals parallel but lie below the soluble $CBOD_5$ removal percentages since the COD test is a measure of the oxygen demand associated with all the carbonaceous compounds found in the wastewater and is indiscriminate as to their biodegradability. However, in the solids residence time operating range for equivalent secondary treatment, there is little biodegradable material left, and, once θ_c's of from 3
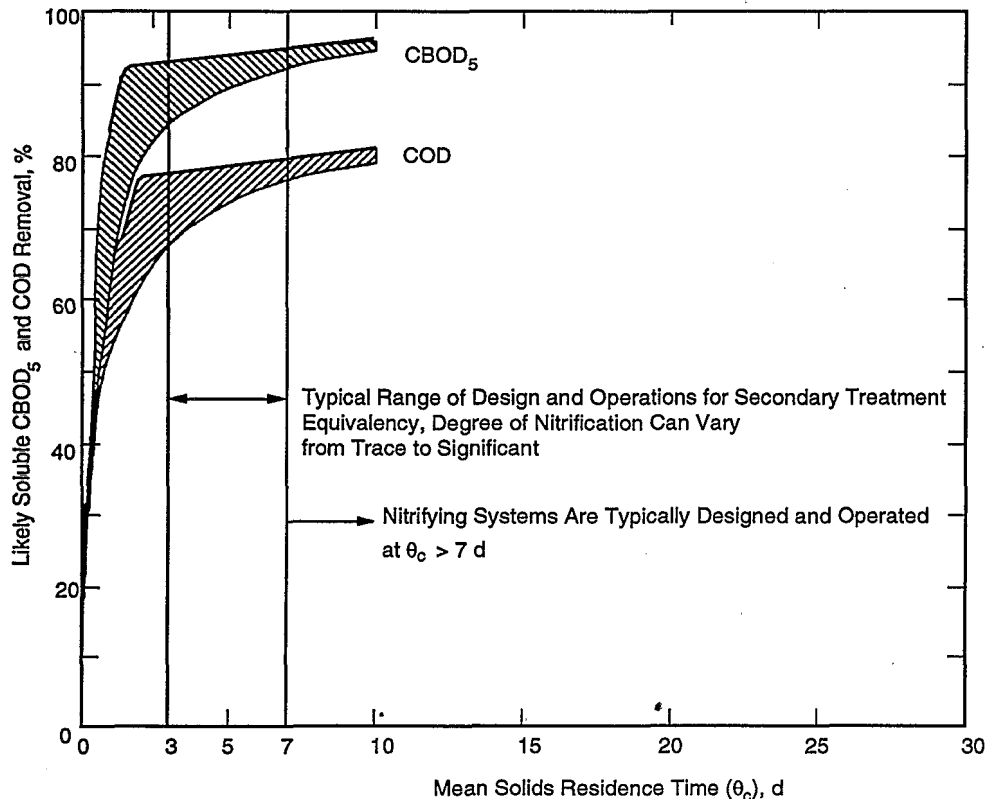
Figure 2-8. Likely soluble $CBOD_5$ and COD removals as a function of mean solids retention time in a biological reactor(s).

to 7 days are encountered, little significant change in the soluble residuals can be expected.

Equipped with the knowledge of the likely soluble substrate ($CBOD_5$ or COD) removals for any solids residence time condition, the next task for the designer is to estimate the solids production derived from the biological reactor. This solids production estimate is partially determined from the inert solids that are applied to and, if applicable, formed in the system (e.g., through biologically enhanced or chemically induced phosphorus removals) and the coefficients that are used to approximate the complex response of the biological reactor as it removes the applied biodegradable and nonbiodegradable substrate. Figure 2-3 and Section 2.6.1.3 fundamentally portray and describe this response.

Table 2-13 presents the basic relationships and provides the coefficients that have been used to describe the solids production characteristics of the biological reactor. This is shown both in terms of the simplistic characterizations generally used in the past, and the more complex relationships that have recently emerged and gained some acceptance in the field, principally because of the development of readily available, conveniently packaged, nonproprietary software and the emergence of powerful personal computers (45,46). (Chapter 5 provides a more in-depth review of the International Association on Water Pollution Research and Control (IAWPRC) model for the interested reader.) The following paragraphs provide a broadly based discussion of these relationships for purposes of the design examples, and allow the reader to formulate some understanding of the significance of any predicted result.

Table 2-13. Volatile Solids Production Considerations and Assumptions

	Classical Approach		IAWPRC Approach
	BOD$_5$ Basis	COD Basis	COD Basis
Basic Relationships	$Y_{net} = \dfrac{Y(\Delta substrate)}{(1 + b\theta_c)}$		$Y_{net} = X_o + \dfrac{Y(\Delta\ substrate)}{(1 + b\theta_c)}\ [1 + (1-f_b)(b\theta_c)]$
Parameters			
Y_{net}	Net VSS production		Net VSS production
Y	Overall Yield Coefficient, VSS/substrate removed		Overall True Yield Coefficient, VSS/biodegradable substrate removed
Δ substrate	Total in minus soluble out		Total in minus soluble out, biodegradable substrate·only
b	Decay coefficient, VSS destroyed/d/reactor VSS		Decay coefficient, VSS destroyed/d/reactor VSS (biodegradable substrate, active biomass basis)
θ_c	Solids residence time, days		Solids residence time, days
X_o	Nonbiodegradable applied VSS, implicitly assumed in definition of coefficients for Y and b		Nonbiodegradable applied VSS
f_b	Biodegradable fraction of active biomass, implicitly assumed in definition of coefficients for Y and b		Biodegradable fraction of active biomass
Typical Coefficients ("Constants") Used for Municipal Wastewaters			
Y	0.7	0.5	0.45
b	0.06	0.1	0.25
X_o			30% of raw VSS*
f_b			0.8

*Should consistently apply the same assumption to the particulate phase COD, TKN, and P. X_o widely varies in reported literature, undoubtedly because of the presence or absence of recycles and collection system influences on the native raw wastewater characteristics. Thirty percent was elected for use in the design example as generally representative of raw municipal wastewaters. Remember that inert nonvolatile solids must be added to compute total solids production.

COD (Substrate) Mass Balance:

$$\left[\begin{array}{c}\text{Total COD}\\\text{(Substrate)}\\\text{Influent}\end{array}\right] = \left[\begin{array}{c}\text{Soluble}\\\text{Effluent}\\\text{COD}\\\text{(Substrate)}\end{array}\right] + \left[\begin{array}{c}\text{COD of}\\\text{Biomass}\\\text{Solids}\\\text{Production}\end{array}\right] + \left[\begin{array}{c}\text{Net COD of}\\\text{Substrate}\\\text{Lost by}\\\text{Oxidation}\end{array}\right]$$
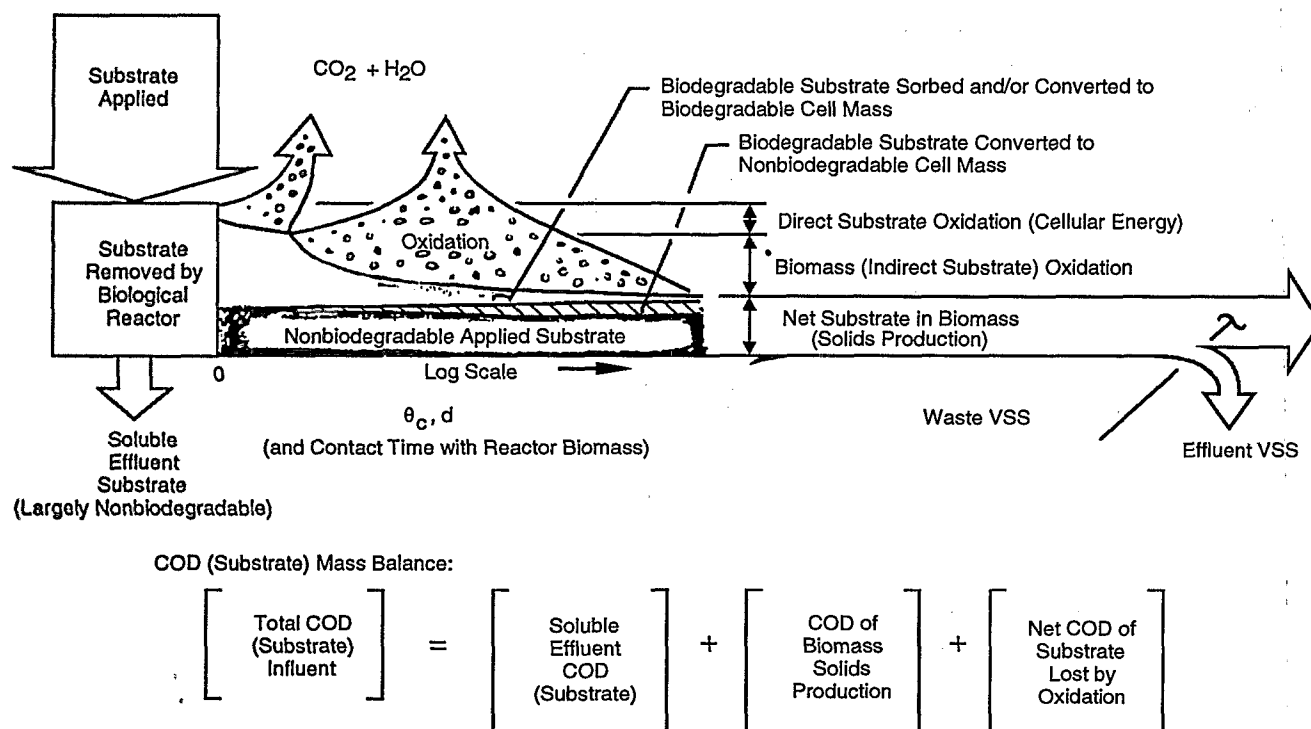
Note :   Theoretical COD of cells is classically taken at 1.42 VSS. Credit for nonbiodegradable
applied substrate will cause the COD of the biomass to increase (likely to 1.5 VSS).
Alternatively, a COD/VSS of 1.4 could be used to develop a safety factor for determination
of the oxygen supply (and realization that the COD test may not fully measure all of the
carbonaceous oxygen demand) or the safety factor could be applied to the end result.

Figure 2-9. Characterization of biological reactor substrate distribution.

Figure 2-9, a simpler version of Figure 2-3, characterizes the distribution of the substrate (COD) upon application to a biological reactor. As explained earlier with Figure 2-8, the soluble substrate is initially removed at a rapid rate and then progressively attenuates to some stable residual which is dependent on the relative biodegradability of the individual organic compounds that compose the substrate and the biological reactor's solids inventory. Older solids residence times allow more acclimation and more soluble substrate removal. The removed particulate substrate sorbs on, and is hydrolyzed prior to transport within the biomass.

Biodegradable material is metabolized (oxidized while providing the energy for cellular replication) and synthesized (into new biomass). The reactions for the applied biodegradable particulate matter are slower than for the soluble substrate since these must be preceded by hydrolysis to reduce the particle size for ready cellular utilization. Nonbiodegradable material, whether inert or volatile, merely comes along for the ride (sometimes masking potentially active sites of microbial stabilization).

The formed or stored cellular products consist of both degradable and nonbiodegradable constituents. The degradable products are oxidized and resynthesized in a never-ending chain of events, leading to the progressive accumulation of nonbiodegradable cellular mass residue. The net solids production of the biological reactor is both substrate and solids residence time dependent. Older solids retention times yield the highest degree of stabilization and the least amount of waste solids for subsequent processing.

With the foregoing understandings, Figure 2-10 was prepared to illustrate the use of the coefficients defined in Table 2-13 with the design example average day raw and settled wastewater characteristics described in Tables 2-11 and 2-14, respectively. The settled wastewater characteristics are based on an assumed 65-percent removal of the raw wastewater SS in the primary clarifiers and 80-percent removal of its nonbiodegradable component. These assumptions along with the assumed nonbiodegradable volatile particulate fraction of 30 percent in the raw wastewater (i.e., nondegradable VSS of 36.9 mg/L) yield a 17-percent nonbiodegradable volatile particulate component (i.e., 7.4 mg/L) in the settled wastewater which contains a total VSS of 43 mg/L; this is consistent with the knowledge that poorer biodegradability tracks larger sized particles. Soluble substrate removals follow
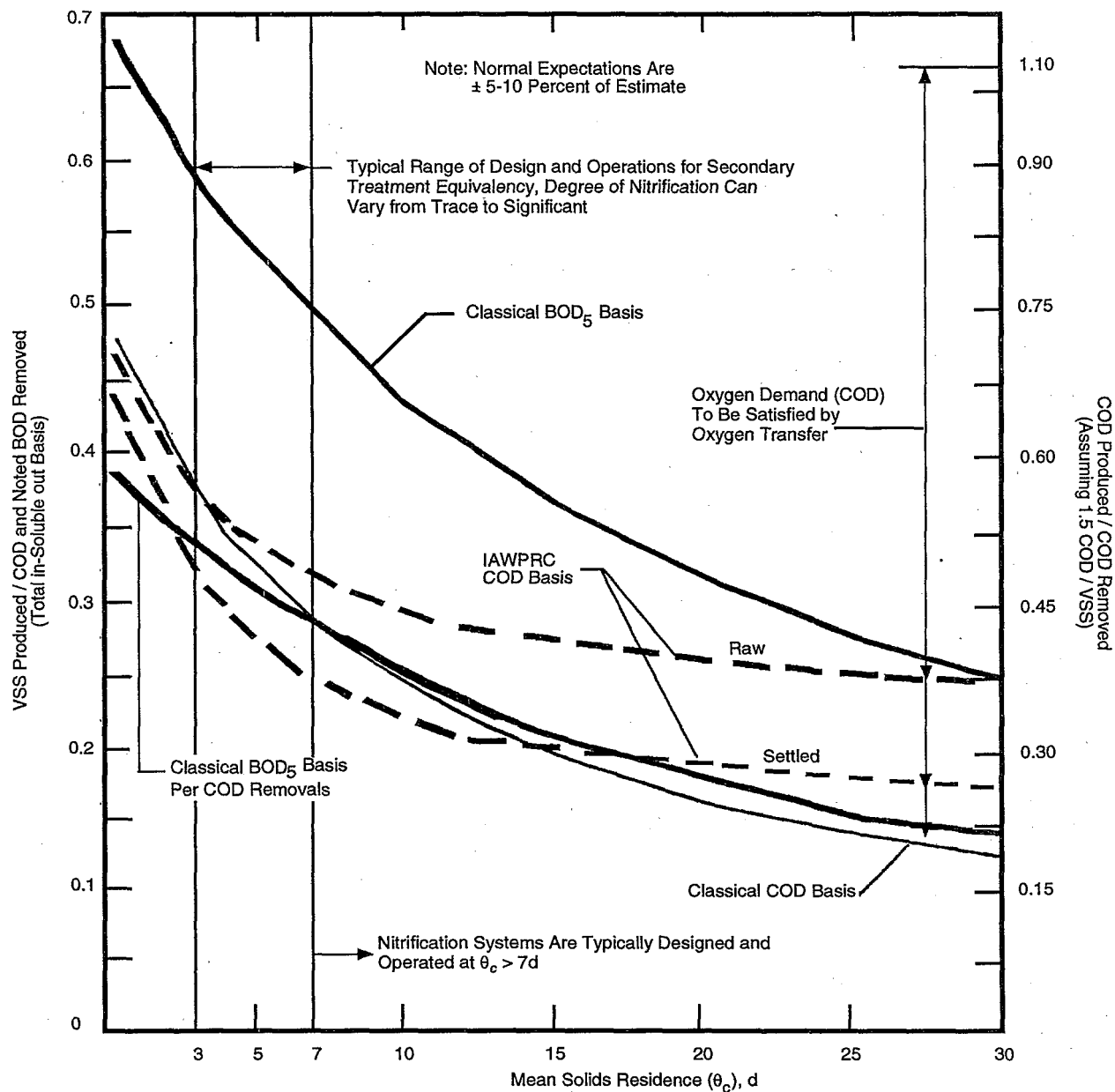
Figure 2-10. Estimates of volatile solids production (see Table 2-11 to Table 2-13 and Figure 2-14).

the estimates contained in Figure 2-8, assuming no additional removal once θ_c exceeds 10 days.

When using the coefficients defined in Table 2-13, it is very important to understand that they are coupled parameters and that they were determined over a particular range of observed operation with specific wastewaters to yield some net prediction of response. Practically, the prediction of the net response is more important than the actual values used for the coefficients. When the coefficients are used outside the operating range from which they were determined, there is an increased opportunity for error. Additionally, coefficient values that have their basis in the classical approach can not be interchanged

with coefficient values that have their basis in the IAWPRC approach.

The coefficients shown in Table 2-13 are often reported with some suggestion of precision and universal applicability. This should not be assumed. In the case of the classical BOD_5 basis of determination, significantly varying and sometimes not necessarily sound parameters have found their way into practice. With the IAWPRC approach, it should be noted that predicted results are strongly influenced by not only the coefficients but also the assumptions used to define the nonbiodegradable component of the particulates in applied wastewater. Although this approach more correctly duplicates theory, its

use still requires utilizing the appropriate model coefficients for the wastewater in question. Presently, the basis for the IAWPRC approach's coefficients represents only a relatively small sampling of wastewaters and treatment plants in the United States, although the Europeans have collected a great deal of data on model parameters. Enthusiasm for any approach must be dampened by the realization that actual solids production values may be strongly influenced by plant specific recycles, and the possible errors that may be imbedded into plant solids production estimates by failure to account for recycle impacts.

The foregoing discussion facilitates understanding of the predicted solids production values found in Figure 2-10. As shown, the coefficients identified for the classical BOD_5 and COD approaches are insensitive to whether the applied wastewater is raw or settled. This, more than anything else, is the great value of the IAWPRC approach, for it allows focus and ready understanding that processing raw wastewaters yields a greater fraction of inert volatile biomass.

Interestingly, the plots in Figure 2-10 show similar predicted results (within 10 percent) through the entire mean solids residence time range of normally encountered designs and operations (θ_c of 3 to 30 days) for the classical BOD_5 and COD basis predictions. Both also fall within the raw and settled sewage envelope defined by the IAWPRC approach over a θ_c of 3 to 15 days. Such would be expected in the classical attempt to define one constant coefficient for raw and settled wastewaters over the range of conventional practice. Undoubtedly, the classical approach parameters were developed and applied largely for secondary treatment equivalency facilities. The stimulant for the IAWPRC approach is found with greater use of older residence time cultures designed to achieve a variety of processing objectives. The IAWPRC approach undoubtedly more fairly characterizes the solids production derived with high θ_c systems; however, informed adjustments of the endogenous decay term in the classical approach would result in the same predictions.

The net conclusion associated with this discussion is that traditional methods are not necessarily bad when properly applied. What is important is not the methods but the predicted result. The mass balances prepared for the design examples use the predictions of the IAWPRC approach to illustrate the tradeoffs between raw and settled wastewater processing.

The oxygen demand that must be satisfied with the biological reactor represents another significant variable that must be estimated by the designer. In contrast to the uncertain (and often arbitrary) application of factors associated with the BOD_5 removal, the oxygen needs of the system readily flow from the COD mass balance with reasonable predictions of the biological reactor's volatile solids production. Figure 2-9 conceptually, and Figure

2-10 precisely, illustrate the linkage and power of a COD mass balance in predicting the carbonaceous oxygen demand with the volatile solids production. Use of a COD mass balance inherently achieves consistency and checks on the predicted carbonaceous oxygen demand and volatile solids production. (Where plants don't routinely perform COD analyses, the designer may be better served by applying routinely anticipated BOD:COD ratios, rather than relying totally on the somewhat arbitrary ratios of oxygen demand to BOD removed or applied.)

When the COD mass balance is applied with a nitrogen mass balance, the nitrogen available for oxidation is readily identified. This technique is superior to merely assuming that the ammonium found in the applied wastewater is what must be oxidized (which carries the implicit assumption and conceptual error that all of the organic nitrogen in the applied wastewater is removed with the solids). The total oxygen demand (TOD) of the system is estimated as the sum of the carbonaceous oxygen demand from the COD mass balance and the ammonia nitrogen available for oxidation (multiplied by 4.6). This total oxygen demand approach is recommended for design, and is illustrated in the mass balances prepared for the design examples.

Table 2-14 summarizes the remaining assumptions used to prepare the mass balances. The performance of the primary treatment system in the more complex Plant B was developed to illustrate the concept that the performance of any given process is influenced by the nature and character of the waste streams applied to it. The assumptions shown in Table 2-14 were used in preparing the mass balances (the pollutant SS characterizations determine the unit process performance for the other pollutants).

Solids production was estimated by the IAWPRC approach described in Table 2-13 (45,46). These estimates reflect raw wastewater solids nonbiodegradability described in Table 2-11 in order to demonstrate concepts of nonbiodegradable volatile matter and buildup (for additional understanding, see Chapter 5). Classical approaches can overestimate volatile solids destruction and air requirements for COD mass balances for high solids residence time systems and when processing raw wastewaters. A solids residence time of 15 days was elected for Plant A to achieve well-stabilized sludge; the same was used for Plant B to allow direct comparison with the simpler plant (without primary clarification). No further decay of recycled VSS from filter backwash and the sludge processing train was assumed because of the advanced stabilization of solids generated from the treatment system and for simplicity.

Anaerobic solids stabilization is included in the more complex Plant B in order to illustrate the impact of digester supernatant. The assumptions shown in Table 2-14 were made to reflect the understanding that biodegradable VSS must be solubilized before stabilization can be-

Table 2-14. Remaining Assumptions for Example Mass Balances

PRIMARY TREATMENT

Process Stream Parameter	% Removal in Primaries
Raw Wastewater TSS	65
Raw Wastewater Nondegradable VSS	80
Supernatant, Filter, Backwash Thickener Overflow SS	50

SECONDARY TREATMENT

Solids Production by IAWPRC approach—See Table 2-13

Solids Residence Time, θ_c = 15 d for both Plants A and B

ANAEROBIC DIGESTION

Waste Stream	% VSS Destroyed	% Applied VSS Remaining as Solubilized VDS
Biodegradable Raw Wastewater VSS in Primary Sludge	85	5
Biodegradable Waste Secondary Solids and Backwash VSS	10	5
All Other Recycled VSS	0	0

SOLIDS THICKENING AND DEWATERING PROCESSES

Two-Stage Anaerobic Digester

95% Solids Capture in Thickening and Dewatering

Separate Primary and Secondary Thickening

gin, and that only the biodegradable material can be stabilized. The advanced stabilization of the secondary solids at the 15-day solids retention time and backwash solids results in little additional stabilization of secondary solids in the anaerobic digester.

Ninety-five percent SS capture was assumed for thickening and dewatering processes. If a gravity thickener was elected (often a good choice for small plants), high-level SS captures are most assured by avoiding excessive blankets in the thickener and using downstream storage. Wash water, as would be encountered if a belt filter was used for dewatering, was ignored in the mass balance.

The example elected not to thicken primary solids separately and not to provide cothickening with secondary solids. This results in direct application of primary sludge for more conservative design. If a gravity thickener was provided, consideration should be given to sizing it for cothickening for additional operating flexibility; although, in this application, odors may result from mixing the raw and secondary solids. Feasible operation might be to base-load the primary sludge into the system, and to have fractional diversions when sludge processing peaks are high. Preferential thickening should always be given to the secondary solids.

A final effluent recycle to the thickener might provide sufficient odor mitigation but would probably result in a higher solids loss. A capability to chlorinate this recycle would provide some measure of protection for sulfide odors. In a nonrural environment, covering the thickener with an odor control system should be considered if cothickening is a routinely planned mode of operation.

The second stage digester (of the more complex Plant B) provides solids concentration (gravity thickening) and storage services. Ninety-percent SS capture efficiency was used to reflect the combined thickening-storage nature of the process and likely difficulty in capturing anaerobically stabilized waste biological sludges.

Federal guidelines for the management of sewage sludge for beneficial agricultural utilization call for a well-stabilized material, free from pathogens, nuisances, and selected hazardous materials. Sewage sludge is classified into two categories, Class A and Class B, based upon the degree of pathogen reduction. Restrictions placed on end uses of sewage sludge are affected by its pathogen reduction classification. Bulk sewage sludge applied to agricultrual and nonagricultural land (e.g., forest, public contact sites, and reclamation sites) must meet at least Class B requirements. Bulk sewage sludge applied to lawns and home gardens, and sewage sludge sold or given away in bags or other containers must meet class A criteria and one of 10 vector attraction mandates. A vector attraction reduction requirement also must be met when sewage sludge is applied to land. Examples of

Class A processes are pasteurization, well-run composting, and alkaline treatment. Class B processes include anaerobic digestion, lime stabilization, and aerobic digestion. Proper control of vectors requires steps such as the following: well-run digestion systems, lime stabilization, drying, and soil incorporation (5).

2.9.3.5 Prepare Mass Balance

The extent of, and need for, process design mass balances depends on the decisions to be made. The processing schematics described earlier in the discussions of Figures 2-5 and 2-6 differ in their approach to raw settleable solids control and stabilization, while remaining insensitive to the specifics of the nitrogen control technology. Conceptually, such an approach allows for the identification of the right nitrogen control technology as a function of the integrated plant works and its processing objectives. For example, if anaerobic digestion was provided without equalization and bleed back of its return, a suspended growth reactor would be preferred over an attached growth reactor because of its long reactor detention time, which would help mitigate the unoxidized nitrogen spikes applied to the reactor.

After determining the conceptual processing sequence of interest, the next step is to characterize the performance of the facility under the conditions of interest to the design issue. Practically, these conditions of interest are best described in terms of some set condition such as the average design day to assure a common reference point.

Tables 2-15 and 2-16 describe the resulting average day mass balance characterization of the simple Plant A and more complex Plant B treatment facilities, respectively. The "mg/L equivalent" term is readily converted to pollutant mass by correcting for the actual flow of the facility and its processing sidestreams. The interrelationship between the "mg/L equivalent" and the actual pollutant concentration is made readily apparent by comparing the equivalent and actual TSS concentrations in Tables 2-15 or 2-16. The use of "mg/L equivalent" as the first step of any plant's characterization allows ready understanding of the interrelationship of processes and pollutants in simple, easy to follow units.

Pollutant mass balances are prepared through the iteration of repetitive calculations. Typically, three iterations are performed through the integrated works to achieve approximate equilibrium values.

Callout 4 in the mass balance for the more complex Plant B represents the reactor effluent quality for a biological reactor with a solids residence time of two days. It is provided because often an existing plant must be upgraded to achieve a new design objective. As shown, the reactor effluent soluble BOD_5 and COD are slightly inferior to that expected from a system with a solids residence time of 15 days. However, lower soluble phosphorus and nitrogen residuals are achieved in the 2-day solids resi-

dence time system because of the higher anticipated degree of sludge synthesis.

Separation of the carbonaceous and nitrogenous oxygen demands naturally occurs in attached growth systems, and is the basis of the two-stage activated sludge system which was applied to several large wastewater treatment plants in the early seventies. Attached growth systems are readily amenable to such concepts since they naturally achieve a staged operation. If the phases are uncoupled through the use of a clarifier, use of a low solids residence time in the first stage, as suggested in Callout 4, offers the opportunity for reactor savings. This concept is addressed in more detail in the suspended and attached growth system design chapters of the manual.

As shown in Figures 2-5 and 2-6, the two plants have a soluble phosphorus residual of about 5 mg/L. Removal of this phosphorus down to a soluble residual of 0.5 mg/L is likely to result in some 15 to 30 mg/L of net inert SS depending on the phosphorus removal strategy (with the range defined by normal expectations for enhanced biological phosphorus removal and immobilization by iron additions). The additional solids mass due to phosphorus removal can be easily quantified at each processing point by adjustment of the inert solids (and immobilized phosphorus) and by the capture efficiency of each unit process. At equilibrium with the capture efficiencies of the processes, the inert SS with phosphorus removal would rise by around 20 percent to yield a total incremental waste solids gain from the liquid processing train of about 18 to 36 mg/L equivalent.

If the phosphorus removal strategy was incorporated into the biological treatment system, the waste secondary solids from Plants A and B would increase by 15 to 30 percent and 35 to 70 percent, respectively. This increase may have significant bearing on the design of suspended growth systems since it would require an equivalent increase in the reactor tankage to maintain the same MLSS concentration as encountered without enhanced phosphorus removal.

When phosphorus removal is anticipated by metal salts, the alkalinity demand must be considered to assure that sufficient alkalinity remains, or is provided, to satisfy the needs of the nitrification reaction. As discussed earlier in Section 2.7.2.1, addition of the metal salts to the reactor effluent, prior to solids separation, is the generally preferred point of application since it allows the often accompanying favorable pH depression to occur without great concern over protection of the nitrification reaction. Here the lowest likely target residual alkalinity is approximately 20 to 30 mg/L as $CaCO_3$ in order to keep the product liquor pH above 6 to 6.5.

2.9.3.6 Use the Mass Balance

The bottoms of Tables 2-15 and 2-16 illustrate the potential use of the information that can be derived from the

Table 2-15. Mass Balance for Plant A (see also Figure 2-5)

| | SS | | | CBOD5 | | | COD | | | Phosphorus as P | | | Nitrogen as N | | | Normalized Flow[b] | |
| | | | | | | | | | | | | | | | | Actual | Flow Fraction |
Processing Point	ISS	VSS	TSS	Sol.	Part.	Tot.	Sol.	Part.	Tot.	Sol.	Part.	Tot.	Sol.	Part.	Tot.	mg/L SS	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	138	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	5,000	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] • Nonbiodegradable 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 sec/m)] = 0.107 m^3/s (or 226.7 cfm)

- and the MLSS in a 18-hour detention time aerator, for a solids residence time of 15 days is

= (15 days)(138 mg/L/d)(24 hr/18 hr) = 2,800 mg/L

• 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

[e] CaO added at 25% of dry weight mass, which, for this example, was assumed satisfactory for PFRP (Processes for Further Reduction of Pathogens) requirements.

Table 2-16. Mass Balance for Plant B (see also Figure 2-6)

Processing Point	SS			CBOD5			COD			Phosphorus as P			Nitrogen as N			Actual mg/L SS	Flow Fraction of Q
	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	5.4	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. Fil. Rx Effl.[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	5,000	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. aft. 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	6,200	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	2,400	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																

Header row spanning "mg/L Equivalents" over SS through Nitrogen as N; "Normalized Flow[a]" over Actual mg/L SS and Flow Fraction of Q.

[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) =   111 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 sec/m)] =0.09 m^3/s (or 191 cfm)
- and the MLSS in an 18-hr detention time aerator, for a solids residence time of 15 days is
= (15 d)(67 mg/L/d)(24 hr/18 hr) = 1,340 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 $CaCO_3$ alkalinity

mass balances. Inspection of the mass balance results illustrates several considerations worthy of note, including the following:

- Design of the nitrogen control system must address the total biodegradable nitrogen delivered to the reactor, not just the ammonium nitrogen.

- The net waste solids found in the dewatered sludge cake and the nitrogen available for oxidation are not substantially different in the two alternatives (because of the commonality of the design objective for solids stabilization).

- The more complex Plant B's use of anaerobic digestion represents a simple processing tradeoff of generating a possible useful end product (methane) against the simpler Plant A's use of oxygenation (and mixing) energy, possibly yielding more optimized reactor volumes and operating costs against more troublesome recycles.

The principal hydraulic recycle of concern at most wastewater treatment plants is that associated with intermittent backwashing of any effluent filtration system (including a denitrification filter). The solids handling operations represent the principal source of pollutant mass-related problems. These problems can be associated with the failure of the solids processing facility to achieve the desired solids capture, and/or the discontinuous nature of sludge wastage and processing practices at any given wastewater treatment plant. In terms of the latter, it is axiomatic in wastewater treatment that as plants decline in size, wastage and sludge processing will become less continuous (as does their monitoring program).

The average day mass balance information defined in Tables 2-15 and 2-16 can not be used directly for design beyond serving as a convenient reference point. It must be adjusted for the controlling peaking factors associated with the minima and maxima elected for design, the anticipated sludge wastage and dewatering operation, and the buffering found in the process train itself. These peaking factors should not be confused with a safety factor, which is applied to reflect uncertainty with the performance of the chosen technology. The following paragraphs describe the thought processes used to develop the controlling design conditions from the average day mass balance.

- Controlling Design Maxima

The controlling design maxima vary as a function of the item under consideration. The controlling condition for the design of the biological reactor can be determined by comparing the ratios of the maximum week to maximum month for the design effluent objectives (1.5 from Table 2-10) to the controlling influent pollutant mass peaking factors used to characterize the influent wastewater (1.3 for the SS and organics from Table 2-12). As is normally encountered, the maximum month is the controlling design condition since the controlling pollutant mass peak-

ing factor is less than the allowable deviation of the effluent standard. Thus, the biological reactor should be designed for successful operation during the maximum month. Since the maximum monthly peaks for the nitrogen mass are less than for the SS and organics, the predicted result of the average day mass balance could be protectively and simply adjusted upward by the maximum month peaking factor to adequately characterize system needs. (Similar thought processes are utilized for effluent standards written in terms of maximum day limitations. Permit writers should take care in establishing maximum day limitations and assure that they are needed since they correspond to an implicit 99.7-percent design reliability objective. Such objectives may result in a substantial increase in the capital cost of any wastewater treatment facility.)

The controlling design maxima for other processing considerations are not the same. The maximum week flow might be used with the reactor's operating level of solids for the maximum month to size the final sedimentation system. Alternatively, it may be necessary to base the design on peak wet-weather flow coupled with an assumption about reactor MLSS. Consider also that reactor MLSS settleability characteristics are variable. The maximum day's flow might be used to size the return sludge system if activated sludge was the preferred technology. The maximum day's peaking factor for the organics and TKN might be used to size an activated sludge system's oxygen supply and dissolution system.

- Controlling Design Minima

The controlling design minima can generally be determined by using the predicted average day mass balance results, adjusted by the minimum month's peaking factors for the lowest-flow year in the facility's design life. This condition is used to develop the minimum needs of the treatment facility. An example would be the minimum air supply that may have to be transferred by the aeration system for an activated sludge plant with a check of mixing under these conditions.

- Anticipated Sludge Wasting and Dewatering Operations

The importance of discontinuous recycles is illustrated in Table 2-17 with the high SS anaerobic digester supernatant and low SS dewatering filtrate described in the mass balance prepared for the more complex Plant B (Table 2-16). As shown for the example, the most stressful conditions at the plant are encountered on the weekends and, at least for the anaerobic digestion characterization of the example, result in a disproportionate recycle of soluble nitrogen.

It is this disproportionate character of the recycles that imposes the greatest stress on the nitrogen control technology—from the standpoint not only of simple reaction kinetics but also of maintenance of sufficient alkalinity to avoid transient pH suppressions and ammonium nitrogen

losses from the treatment plant. Clearly, as operations become more discontinuous, any attempt to optimize the liquid processing train incorporates greater risk because of the loads associated with the recycles. These loads may, and often do, have peaks which are far more frequent and severe than anything encountered in the raw wastewater.

Mitigation measures for these transients include equalization and bleedback, and avoidance of plug flow, short contact time reactors. With attached growth systems, if anaerobic digestion is elected for solids stabilization, the designer would be well advised to use equalization and bleedback of supernatant and filtrate to avoid sudden soluble nitrogen loads from the solids processing train to the short hydraulic contact time reactors.

• Layered Diurnal Peaks

Wastewater flows and loads are not constant during the operating day. One-half or more of the plant's daily load may well arrive over an eight-hour period. This corresponds to an eight-hour diurnal peaking factor of 1.5 times the average daily value. Maximum hourly loads often are more than two times the average daily value. In general, the smaller the plant (with its smaller collection system), the more severe the diurnal peaking factor.

There are no hard and fast rules for election of the proper peaking factor for the raw wastewater loads; they are largely dependent on the elected process train. The liquid volume found in the plant, along with the plant's liquid stream recycles and returns, acts to mitigate the peaks encountered in the raw wastewater and the recycle streams. If adequate aeration capacity is available, the same size tank will produce lower effluent ammonia levels if it is plug flow rather than complete mix in a dynamic load situation. Mitigation of processing peaks is also achieved through the use of complete mix reactors. Further, it should be remembered that some excursions over the effluent standard are allowable.

The elected process peak for a short detention time attached growth reactor should be higher than that selected for a longer detention time suspended growth system. Given that this manual is intended to serve the needs of the designer of the smaller plant and to characterize the different needs of the suspended growth and attached growth technologies, a diurnal wastewater processing peak of 1.3 and 1.6 was elected for the respective technologies. The corresponding diurnal minima for both processes, which should be applied to the minimum month when natural denitrification is anticipated, are about 0.7 and 0.6, respectively. These peaking factors should be applied as appropriate in the design, but, at minimum, they must be applied to the nitrogenous component available for oxidation and denitrification.

The remaining processing peak that should be identified and checked is consideration of the recycles from the solids processing train. The analyses in Table 2-17 suggest that the worst-case condition results in a periodic 60-percent transient two-hour increase in the plant's average day soluble nitrogen load, twice a day. This is the same peaking factor as elected for the attached growth technologies and some 20 percent higher than the peaking factor elected for the suspended growth technologies. Accordingly, no adjustment need be made to the available nitrogenous component for the attached growth technology application. (If the transient solids processing recycle peak was less than the diurnal wastewater peak, the correct methodology would end up with a lower overall combined peak for the mainstream process design.) For convenience with the design examples, and given that the suspended growth system inherently offers greater buffer of applied peaks through its greater liquid volume, the decision was made to apply the elected process diurnal peak also to the total available nitrogen.

• A Word About Dilution, and Lack Thereof

The designer should remember that regulatory effluent standards apply to the lesser of the permitted concentration or mass at the rated average annual daily flow. Under dry weather or draught conditions (or an extremely tight collection system), the favorable soluble pollutant dilution in complying with concentration dependent effluent standards will be lost and noncompliance may be encountered when compliance would be predicted under the elevated flow regimes of the average day or maximum month. Process selections that just barely comply at average day flows may well fail under seasonal minima. A check should be made under low flow conditions to ensure compliance. Fortunately, low flows are encountered with warmer temperatures which, for the suspended growth technologies, gives some reasonable assurance of compensating increases in reaction rates.

A low flow condition check is particularly relevant with attached growth technologies as their performance is largely concentration dependent under oxygen transfer limiting conditions, as explained in Section 2.6.1. Warmer temperatures and lower flows yield lower oxygen transfer to a more concentrated waste stream with a longer residence time. This may yield conditions where poorer performance on both a mass and concentration basis may be encountered in the summer, as opposed to the classical expectation of poorer performance in the winter. This issue is discussed again in Section 2.7 and more fully in the later sections of this manual dealing with attached growth technologies.

Table 2-18, based on the previous paragraphs, summarizes the controlling design conditions for the design examples.

82

Table 2-17. More Complex Plant B Solids Processing Recycle Impacts on Main Processing Stream (see also Table 2-16 for basis of average day condition)

PROCESSING STREAM CHARACTERIZATION IN LIQUID PROCESSING TRAIN, mg/L unless noted otherwise

Processing Point	SS	CBOD$_5$	COD	P	TKN
Raw Influent, total	175	150	280	6.0	30.0
soluble	—	50	100	5.0	24.0

Case 1: Continuous Sludge Processing (or Average Day Condition)

	SS	CBOD$_5$	COD	P	TKN
Supernatant, Factor = 1.0	13	2	13	0.3	1.6
Filtrate, Factor = 1.0	6	2	8	0.2	1.5
Total	19	4	21	0.5	3.1
% Increase over Raw Influent	11	3	8	8.0	10.0

Case 2: Sludge Wasting for Two Hours per Day, Twice a Day, and Dewatering for Six Hours per Day, Five Days per Week

A. Weekdays, Worst Condition: Sludge Wastage to Digester for Two Hours and Concurrent Sludge Dewatering

	SS	CBOD$_5$	COD	P	TKN
Supernatant, Factor = 24/4 = 6	78	12	78	1.8	9.6
Filtrate, Factor = (24/6)(7/5) = 5.6	34	11	45	1.1	8.4
Total	112	23	123	2.9	18.0
% Increase over Raw Influent for the Two, 2-Hour Periods	64	15	44	48	60

B. Weekends, Worst Condition: Sludge Wastage to Digester for Two Hours per Day, Twice a Day, with Full Digester

Supernatant, Factor = (24/4)(0.00509/0.00211) = 14.5

	SS	CBOD$_5$	COD	P	TKN
Total	189	29	189	4.4	23.2
% Increase over Raw Influent for the Two, 2-Hour Periods	108	19	67	73	77
Total Soluble	—	14	44	2.9	16.0
% Increase over Raw Soluble Influent for the Two, 2-Hour periods	0	28	44	58	67

Table 2-18. Summary of Controlling Design Conditions for Biological Reactor with Design Examples

SUMMARY

- Size Biological Reactor Based on Maximum Month Loads
  - —Peaking Factor of 1.3 times average day mass balance for SS, BOD$_5$, and COD (per Table 2-12)
  - —Peaking Factor of 1.2 times average day mass balance for Total Phosphorus and Nitrogen (per Table 2-12)
    - Further adjust available Nitrogen by additional Diurnal Peaking Factor in plant as follows:

	Suspended Growth	Attached Growth
Diurnal Process Peaking Factor	1.30	1.60
Total Processing Peaking Factor	(1.2)(1.3) = 1.56	(1.2)(1.6) = 1.92

  Note: With attached growth technologies, it may be appropriate to apply the SS and Organic Peaking Factor because of oxygen-demanding conditions.
    The Total Processing Peak for these pollutants is (1.3)(1.6) = 2.08
  - —At a minimum, check for concentration compliance at minimum month flow conditions (Peaking Factor = 0.7) with maximum month load.
    - Attached growth technologies also should check for oxygen transfer under these conditions.
- Size Oxygen Supply for Suspended Growth Technologies Based on Maximum Day Peaking Factors of 2.1 for Organics and 1.7 for Available Nitrogen (Table 2-12).
- Size Clarifier for ≥Maximum Week Flows (Peaking Factor = 1.9) with Maximum Month's Operating Solids.

PLANT CHARACTERIZATION AND WASTEWATER CHARACTERISTICS, mg/L Equivalent[a] and Actual[b]

	SIMPLER PLANT A (per Figure 2-5 and Table 2-15)				MORE COMPLEX PLANT B (per Figure 2-6 and Table 2-16)				
		Maximum Month and Noted				Maximum Month and Noted			
	Avg. Day[a]	Process Peak[a]	Diurnal Peak[a]	Low Q Check[b]	Avg. Day[a]	Process Peak[a]	Diurnal Peak[a]		Low Q Check[b]
			Sus. Growth				Sus. Growth	Att. Growth	
Q	1.0	1.0	1.0	0.7	1.0	1.0	1.0		0.7
Process Infl., mg/L									
SS	187	243		347	80	104			148
BOD$_5$	152	198		283	97	126			180
SBOD$_5$	50	65		93	53	69		110	99
COD	290	377		538	187	243			347
SCOD	100	130		186	106	138		221	197
Reactor Effluent[c]									
SS	138	179		256	67	87			124
COD	137	178		254	79	103			147
Available N[d]	24	29	38	41	26	31	41	50	44
Carbon O$_2$ Demand[e]	153	321			108	227			
Nitrogen O$_2$ Demand[d,e]	113	192			120	204			
Total O$_2$ Demand[e]	266	513			228	431			
Wastewater Alkalinity	120	132	132+	188	120	132	132+	132+	188

[a]Average daily flow (ADF) basis, actual lb/day mass is determined by multiplying the designated mg/L Equivalent by the ADF (mgd) and 8.34.
[b]The mg/L value is value expected if the maximum month pollutant load (mass) was experienced during the minimum month flow.
[c]For the special intermediate condition (Callout 4) defined in the average day mass balance for Plant B, the Process Peak values for the Reactor Effluent are as follows: SS = 108, COD = 144, Available N = 30 mg/L Equivalents.
[d]Rounded Available N to nearest whole number and determined O$_2$ demand by directly multiplying by 4.6, ignoring soluble effluent refractory organic N and NH$_4$-N.
[e]Process Peak demands reflect maximum day, not maximum month.

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National Technical Information Service
5285 Port Royal Road
Springfield, VA 22161
703-487-4650

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Chapter 3
Process Chemistry and Kinetics of Biological Nitrification

3.1 Introduction

This chapter presents a review of the process chemistry and kinetics of biological nitrification for wastewater treatment. An understanding of the process fundamentals is important for appreciating the factors affecting the design, operation, and performance of nitrification process systems. In discussing kinetics, the emphasis in this chapter is on defining the intrinsic effects of such factors as ammonia concentration, temperature, and pH. This information is intended to serve as the conceptual basis for Chapter 6, which addresses design aspects of specific nitrification process systems and the influence of such kinetic factors on design.

3.2 Fundamentals of Nitrification

The nitrification process is carried out by bacterial populations that sequentially oxidize ammonium to nitrate with intermediate formation of nitrite. The two principal genera of importance for carrying out this process are *Nitrosomonas* and *Nitrobacter*. Both of these groups are classified as autotrophic organisms because they derive energy for growth from the oxidation of inorganic nitrogen compounds. In contrast, heterotrophic bacteria derive energy from the oxidation of organic matter. Another feature of these organisms is that they use inorganic carbon (carbon dioxide) for synthesis rather than organic carbon. The two groups are distinguished from one another by their ability to oxidize only specific species of nitrogen compounds. While *Nitrosomonas* can oxidize ammonium to nitrite but cannot complete the oxidation to nitrate, *Nitrobacter* is limited to the oxidation of nitrite to nitrate. Since complete nitrification is a sequential reaction, treatment process systems must be designed to provide an environment suitable for the growth of both groups of nitrifying bacteria.

3.2.1 Metabolism and Stoichiometry

On a biochemical level the nitrification process involves more than the sequential oxidation of ammonia to nitrite by *Nitrosomonas* and nitrite to nitrate by *Nitrobacter*. Various reaction intermediates and enzymes are involved (1). Rather than discuss these pathways, however, this manual focuses on the response of the nitrification organisms to environmental conditions. This information is important

for engineering nitrification process systems that ensure that the resident nitrifiers are able to carry out their metabolic activities efficiently.

The stoichiometric equation for the oxidation of ammonium to nitrite by *Nitrosomonas* is:

$$NH_4^+ + 1.5\ O_2 \rightarrow 2H^+ + H_2O + NO_2^- \qquad (3\text{-}1)$$
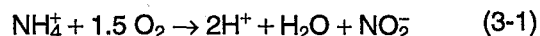
The release of free energy by this reaction at conditions within the cell has been estimated by various researchers to be between 58 and 84 kcal/mole of ammonium (1,2). The reaction for the oxidation of nitrite to nitrate by *Nitrobacter* is:

$$NO_2^- + 0.5\ O_2 \rightarrow NO_3^- \qquad (3\text{-}2)$$

This reaction has been estimated to release 15.4–20.9 kcal/mole of nitrite at conditions typically found within microbial cells (2). Thus, *Nitrosomonas* obtains more energy/mole of nitrogen oxidized than *Nitrobacter*. If it is assumed that the amount of cell mass produced is proportional to the degree of energy release, there should be a greater mass of *Nitrosomonas* formed than *Nitrobacter* per mole of nitrogen oxidized. This is indeed the case, as will be discussed.

The expression for overall oxidation of ammonium by both groups is obtained by adding Equations 3-1 and 3-2:

$$NH_4^+ + 2\ O_2 \rightarrow NO_3^- + 2\ H^+ + H_2O \qquad (3\text{-}3)$$
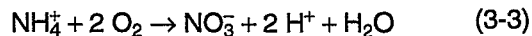
The equations for the synthesis of *Nitrosomonas* and *Nitrobacter* are shown in Equations 3-4 and 3-5, respectively. These assume that the empirical formulation of bacterial cells is $C_5H_7NO_2$:

$$13\ NH_4^+ + 15\ CO_2 \rightarrow 10\ NO_2^- + 3\ C_5H_7NO_2 + 23\ H^+ + 4\ H_2O$$
$$\textit{Nitrosomonas} \qquad (3\text{-}4)$$

$$10\ NO_2^- + 5\ CO_2 + NH_4^+ + 2\ H_2O \rightarrow 10\ NO_3^- + C_5H_7NO_2 + H^+$$
$$\textit{Nitrobacter} \qquad (3\text{-}5)$$
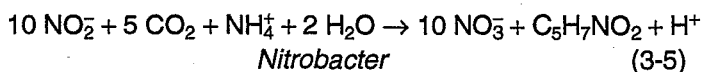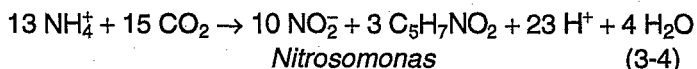
Bacterial cells grow by coupling the reactions that produce energy (Equations 3-1 and 3-2) with those involving cell synthesis (Equations 3-4 and 3-5). Thus, cell synthesis can be described by combining the equations for energy

yield and cell synthesis. The efficiency of the organisms in converting the released energy into biomass dictates how these equations are combined. Efficiency can be measured in terms of the observed yield, expressed as the cell mass produced/mass of substrate utilized. This yield coefficient is normally specified as the mass of VSS produced/mass of ammonium or nitrite oxidized.

Yield values calculated from theoretical energy release relationships are 0.29 g VSS/g of NH_4^+-N and 0.084 g VSS/g of NO_2^--N (2). Yield values observed in experimentation are lower: for the oxidation of ammonium to nitrite by Nitrosomonas, they are 0.04–0.13 g VSS/g NH_4^+-N; and for the oxidation of nitrite to nitrate by Nitrobacter 0.02–0.07 g VSS/g NO_2^--N (1). Observed yields may be lower than theoretical yields because a fraction of the free energy released by oxidation is diverted to microbial maintenance functions. The total yield of nitrifiers, when considering nitrification as a single-step process from ammonium to nitrate, is 0.06–0.20 g VSS/g NH_4^+-N oxidized.

The observed yield will vary with changing environmental conditions and with changes in the rate of growth of the microbial cells, in part accounting for the range of observed yield values. Observed yield is the net yield of microbial cells, which takes into account the process of endogenous decay. The effect of endogenous decay on net yield, however, is typically not considered significant, given the uncertainty in estimating actual Nitrosomonas yields. As such, the net yield coefficient for Nitrosomonas is often considered an estimate of the true yield coefficient (3).

Equations for synthesis of Nitrosomonas and Nitrobacter, using yields of 0.08 g VSS/g NH_4^+-N and 0.05 g VSS/g NO_2^--N, respectively, are as follows:

$$1.00 \ NH_4^+ + 1.44 \ O_2 + 0.0496CO_2 \rightarrow$$
$$0.01 \ C_5H_7O_2N + 0.990 \ NO_2^- +$$
$$0.970 \ H_2O + 1.99 \ H^+ \qquad (3\text{-}6)$$

$$1.00 \ NO_2^- + 0.00619 \ NH_4^+ + 0.031 \ CO_2 +$$
$$0.0124 \ H_2O + 0.50 \ O_2 \rightarrow$$
$$0.00619 \ C_5H_7O_2N + 1.00 \ NO_3^- + 0.00619 \ H^+ \quad (3\text{-}7)$$
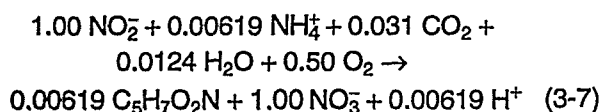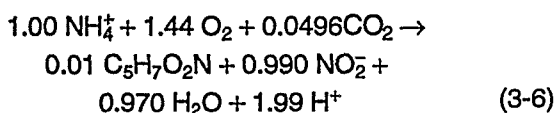
Combining these equations, the overall reaction describing complete nitrification is:

$$1.00 \ NH_4^+ + 1.89 \ O_2 + 0.0805 \ CO_2 \rightarrow$$
$$0.0161 \ C_5H_7O_2N + 0.952 \ H_2O + 0.984$$
$$NO_3^- + 1.98 \ H^+ \qquad (3\text{-}8)$$
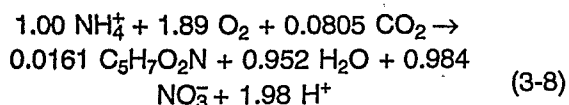
The implications of Equation 3-8 on the design of nitrification systems are significant. The stoichiometric coefficients imply that per mole of ammonium removed the nitrification process requires a significant amount of oxygen, produces a small amount of biomass, and results in substantial destruction of alkalinity through the production of hydrogen ions. For example, according to Equation

3-8, synthesis and oxidation of 20 mg/L ammonium-nitrogen (the equivalent of 25.7 mg/L ammonium) would result in the consumption of 86.4 mg/L oxygen, the production of 2.6 mg/L nitrifying organisms, and the destruction of 141.4 mg/L alkalinity (as $CaCO_3$). Values for oxygen utilization, biomass yield and alkalinity destruction coefficients that are generally accepted in practice for designing nitrification systems are listed in Table 3-1. Note that the oxygen utilization coefficient of 4.6 is conservative in that it reflects the energy reaction only (Equation 3-3) and does not consider N used for cell synthesis.

Table 3-1. Oxygen Utilization, Biomass Yield, and Alkalinity Destruction Coefficients Acceptable for Design of Nitrification Systems

Parameter	Equation	Coefficient
Oxygen utilization	$\dfrac{g \ O_2 \ required}{g \ NH_4^+\text{-}N}$	4.6
Biomass yield	$\dfrac{g \ VSS \ produced \ (as \ nitrifiers)}{g \ NH_4^+\text{-}N}$	0.1
Alkalinity destroyed	$\dfrac{g \ alkalinity \ (as \ CaCO_3)}{g \ NH_4^+\text{-}N}$	7.1

It is important to understand that in virtually all nitrification systems treating municipal wastewaters, biodegradable compounds other than ammonium are present. These compounds will have an influence on total oxygen utilization, biomass production, and alkalinity destruction.

Nitrification reactions take place in an aqueous environment. Thus, the production of free acid (H^+) and the consumption of gaseous carbon dioxide (CO_2), as described by Equations 3-4 and 3-5, will impact the aqueous carbonic acid system equilibria (4). This, in turn, can affect the nitrification reactor pH. As will be discussed in Section 3.3.4, pH affects the growth rate of the nitrifiers. As a result, engineering decisions on the selection of a nitrification reactor system (e.g., pure oxygen versus air-based systems) will influence the resulting pH conditions.

3.3 Nitrification Kinetics

In the context of the nitrification reaction equations that have been presented, kinetics can be considered as the study of the factors influencing the rates of these reactions, and as explanations for these rates (5). Ammonium removal in the nitrification process occurs through microbial synthesis, or growth, and oxidation, according to Equations 3-6 and 3-7. In this section, kinetic expressions will be developed to describe the rate of nitrifier growth and ammonium oxidation, and the impact that a number of environmental factors have on these rates will be considered. Other factors that impact the efficiency and per-

formance of nitrification reactors also will be discussed, including the feed organic carbon (CBOD) to nitrogen ratio, diffusional limitations, and the influence of reductive zones. Although reference will be made to data derived from operating systems, descriptions of these systems as well as information pertaining to their design and performance are presented in Chapter 6.

3.3.1 Kinetics of Biomass Growth and Ammonia Utilization

A description of ammonium and nitrite oxidation can be derived from an examination of the growth kinetics of *Nitrosomonas* and *Nitrobacter*. *Nitrosomonas* growth is limited by the concentration of ammonium, while *Nitrobacter* growth is limited by the concentration of nitrite. The kinetic equation proposed by Monod (6) is used to describe the kinetics of biological growth of either *Nitrosomonas* or *Nitrobacter*:

$$\mu = \hat{\mu}\frac{S}{K_s + S} \qquad (3\text{-}9)$$

where:

μ = specific growth rate of microorganisms, d^{-1}
$\hat{\mu}$ = maximum specific growth rate of microorganisms, d^{-1}
K_s = half-saturation or half-velocity coefficient (equivalent to the growth-limiting substrate concentration at half the maximum specific growth rate), mg/L
S = growth-limiting substrate concentration, mg/L

Nitrite normally does not accumulate in large amounts in biological treatment systems under steady-state conditions. This is because the maximum growth rate of *Nitrobacter* is considerably higher than the maximum growth rate of *Nitrosomonas* and K_s values for both organisms are less than 1 mg/L N at temperatures below 20°C (68°F). For this reason, the rate of nitrifier growth can be modeled with Equation 3-9, using the conversion of ammonium to nitrite as the rate-limiting step:

$$\mu_N = \hat{\mu}_N\frac{N}{K_N + N} \qquad (3\text{-}10)$$

where:

μ_N = specific growth rate of *Nitrosomonas*, d^{-1}
$\hat{\mu}_N$ = maximum specific growth rate of *Nitrosomonas*, d^{-1}
K_N = half-saturation coefficient for *Nitrosomonas*, mg/L NH_4^+-N
N = NH_4^+-N concentration, mg/L

Although the Monod expression is the most widely accepted approach for describing microbial growth kinetics and is acceptable for practical engineering design, it has certain theoretical deficiencies. These are particularly apparent when the expression is used to describe processes that may involve multiple substrate-limiting conditions (e.g., microbial growth limited by ammonium or oxygen under transient versus steady state conditions) and associated multiple organism groups. It is important to recognize the shortcomings of the Monod expression in applications of nitrification kinetics since mass-transport or diffusional resistances, heterotrophic/nitrifier competition, and transient conditions (7) can sometimes negate the assumption that ammonium conversion to nitrite is the rate-limiting step in the nitrification process.

The rate of ammonium oxidation is controlled by the growth of *Nitrosomonas* and is related to this growth by the *Nitrosomonas* yield coefficient. The relationship between the oxidation rate and the growth rate of *Nitrosomonas* can be expressed as follows:

$$q_N = \frac{\mu_N}{Y_N} = \hat{q}_N\frac{N}{K_N + N} \qquad (3\text{-}11)$$

where:

q_N = ammonium oxidation rate, g NH_4^+-N oxidized/g VSS/d
$\hat{q}_N$ = maximum ammonium oxidation rate, g NH_4^+-N oxidized/g VSS/d
Y_N = organism yield coefficient, g *Nitrosomonas* grown (VSS)/g NH_4^+-N removed

The growth of microorganisms may be expressed in terms of their doubling, or generation, time. Generation times of heterotrophic bacteria, those responsible for carbonaceous oxidation or CBOD removal, are normally reported at 10–20 times less than the generation times for nitrifiers (8). Because of the slow growth rate of nitrifiers, a sufficient solids retention time (also referred to as the mean cell residence time or sludge age) is essential in nitrification process systems in order to retain an adequate population of these organisms. The solids retention time in a biological system is normally defined as:

$$\theta_c = \text{(total mass of biological solids)} \div \text{(total mass of biological solids leaving the system/d)} \qquad (3\text{-}12)$$

where:

θ_c = the solids retention time (or sludge age or mean cell residence time), d

At steady state, the solids leaving the system will be equal to the solids produced. Therefore, the growth rate and solids retention time of the organisms in the system are related by:

$$\frac{1}{\theta_c} = \mu_N - b_N = \mu'_N \qquad (3\text{-}13)$$

where:

μ'_N = net specific growth rate of nitrifiers, d^{-1}

b_N = endogenous decay coefficient for nitrifiers, d^{-1}

With nitrifying organisms, b_N is often considered to be negligible (i.e., $b_N = 0$), in which case the specific growth rate, μ_N, is the same as the net specific growth rate, μ'_N.

Values for the maximum specific *Nitrosomonas* growth rate and the corresponding half-saturation coefficient are presented in Table 3-2. These are typical of those reported in the literature. The values of $\hat{\mu}_N$ are at least an order of magnitude smaller than typical $\hat{\mu}$ values for heterotrophs, implying the need for a much longer solids retention time for biological systems designed to achieve nitrification versus only carbon oxidation. The values of K_N presented in Table 3-2, although quite low, exceed values reported elsewhere (4).

Table 3-2. Maximum Specific Growth Rates and Half-Saturation Coefficient Values for *Nitrosomonas* at Constant Temperature (20°C) (Adapted from Reference 9)

$\hat{\mu}_N,$ d^{-1}	$K_N,$ mg/L NH_4^+-N	Reference
1.32	3.6	10
0.84	1.0	3
1.62	0.6	11

The significance of low K_N values is clear from examination of Equations 3-10 and 3-11. When K_N is low with respect to N, the growth rate and the ammonium oxidation rate are independent of the concentration of ammonium and the *Nitrosomonas* organisms are growing at their maximum rate. In complete mix activated sludge systems, however, N can be lower than K_N, in which case kinetics approach first order (i.e., growth rate is dependent on substrate concentration). The independence of growth rate from substrate concentration, characterized as *zero-order kinetics*, has been observed by a number of researchers (12–15).

The maximum specific growth rate coefficient of *Nitrosomonas* is highly dependent on the constituents in the wastewater and should be determined experimentally, particularly when treating an industrial wastewater or a municipal wastewater with a significant industrial input. A simple bench scale, laboratory procedure for determining $\hat{\mu}_N$ is described in detail elsewhere (16).

A number of environmental factors significantly influence nitrifier growth rates, thus impacting the minimum cell residence time required to ensure sufficient buildup and retention of nitrifiers in a biological system. While factors affecting process kinetics may not influence the intrinsic nitrifier growth rates, they will affect the selection of other process design parameter values. For example, mass transport or diffusional resistances in attached growth reactors will increase the required reactor solids retention time as the nitrifiers are no longer operating at their intrinsic growth rates. The effect of such factors on nitrification kinetics, together with the effect of environmental factors, will be considered in the following sections.

3.3.2 Temperature Effects

The nitrification process occurs over a range of approximately 4–45°C (39–113°F), with about 35°C (95°F) optimum for *Nitrosomonas* (17) and 35–42°C (95–108°F) optimum for *Nitrobacter* (18,19). The process has been shown to be strongly dependent on temperature. Quantifying the temperature effect with confidence is difficult, as demonstrated by the widely reported observations in the literature. The collection of conclusive data is complicated, in part, by the fact that both the maximum growth rate and the half-velocity coefficients of nitrification are temperature sensitive (4).

Conservative estimates for the maximum growth rate of *Nitrosomonas* over the temperature range of 10–30°C (50–86°F) are presented in Table 3-3 (20). The fact that the 20°C (68°F) value for $\hat{\mu}_N$ in Table 3-3 is less than those presented in Table 3-2 simply illustrates the variation in rates reported in the literature.

Table 3-3. Maximum Specific Growth Rate Values for *Nitrosomonas* as a Function of Temperature

Temperature, °C	$\hat{\mu}_N,$ d^{-1}
10	0.3
20	0.65
30	1.2

The values in Table 3-3 agree reasonably with the van't Hoff-Arrhenius equation, which predicts the doubling of growth rates with each 10°C increment in temperature (20). Arrhenius-type relationships have been observed by a number of researchers who have measured the oxidation rate of *Nitrosomonas* as a function of temperature in various environments over the 5–30°C (41–86°F) range. The nitrification rate has been observed to decrease above 30–35°C (86–95°F) (4). This apparent optimum temperature range is the result of two interactive processes: the anticipated increase in reaction rate with increase in temperature and protein denaturation above a critical temperature. For design purposes, an acceptable Arrhenius-type expression of the effect of temperature on the maximum growth rate of *Nitrosomonas* over a temperature range of 5–30°C (41–86°F) is:

$$\hat{\mu}_N = 0.47\ e^{0.098\ (T-15)} \tag{3-14}$$

90

where:

T = temperature, °C

This is graphically displayed on Figure 3-1. Originally presented in the 1975 edition of this manual (4), subsequent studies (23–27) have tended to confirm the expression. Although other expressions have been cited (4,20,22) and have been found to be as acceptable, Equation 3-14 has found some consensus among designers and is used as the default value in the International Association on Water Pollution Research and Control (IAWPRC) model for suspended growth process design (see Chapter 5). Although K_N has also been reported to vary according to an Arrhenius-type relationship (11), the low value of the coefficient and the reported range of values—even at a constant temperature (Table 3-2)—imply that selecting a constant value of 1.0 mg/L NH_4^+-N should be acceptable for design purposes.

The variation in $\hat{\mu}_N$ and K_N in the literature (even at constant temperature) may be related in part to reactor biomass concentration differences, according to the work of Shammas (21). Using biomass derived from identical fill-and-draw, completely mixed activated sludge reactors, Shammas concluded that there is an interaction between biomass concentration, temperature, and pH. Consequently, he developed a relationship that expresses temperature sensitivity as a function of the reactor biomass concentration.

3.3.3 Effect of DO Concentration

The concentration of DO has a significant effect on the rates of nitrifier growth and nitrification in biological waste treatment systems. By modeling the growth of *Nitrosomonas* according to the Monod equation (Equation 3-9), with DO as the growth-limiting substrate concentration, values for the half-saturation coefficient have been reported as 0.15–2.0 mg/L O_2 (4). Evidence suggests that the value for the coefficient increases with increasing temperature (28).

Historically, the influence of DO on nitrification rates has been controversial. Qualitative observations imply that under certain conditions complete nitrification can be achieved in biological systems at DO levels as low as 0.5 mg/L (4). A recent comprehensive study sought to provide a clearer quantification of the effects of DO on nitrification and to identify interdependent factors affecting the relationship (7). The following can be implied from the results of the study, when defining the relationship between DO and nitrification kinetics:

- The value of DO at which nitrification is limited can be 0.5-2.5 mg/L in either suspended or attached growth systems under steady state conditions, depending on the degree of mass-transport or diffusional resistances and the solids retention time.

- A high solids retention time may be required to ensure complete nitrification at low DO concentrations, and for conditions where diffusional resistances are significant.
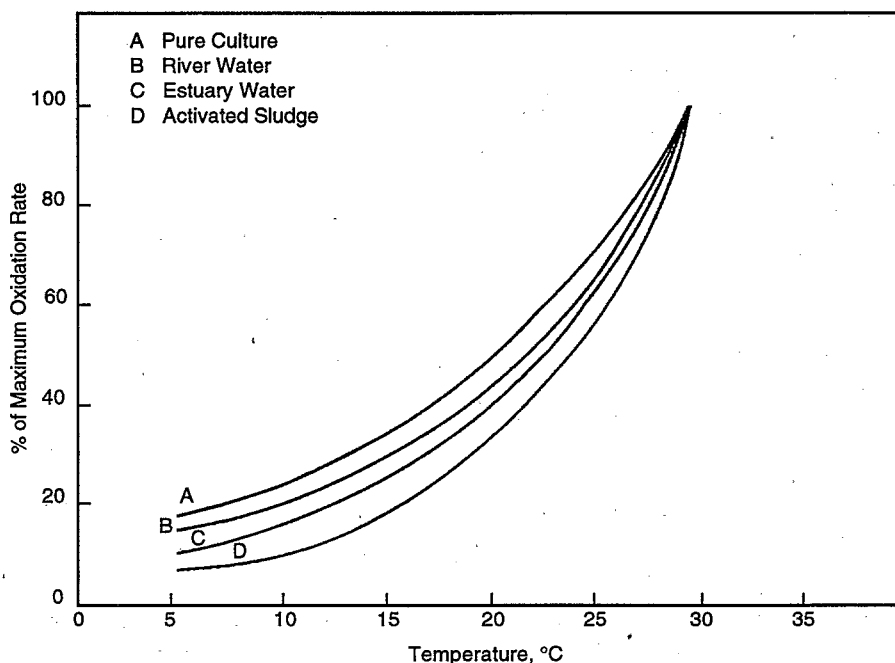
Figure 3-1. Effect of temperature on oxidation of ammonium by *Nitrosomonas* (adapted from Reference 21).

- Under transient conditions of organic shock loading, diffusional resistances and heterotrophic/nitrifier competition can increase the limiting DO value significantly.

- Under transient conditions, nitrite conversion to nitrate can become the rate-limiting step in the nitrification process; in such conditions, the resulting accumulation of nitrite is not correlated to low DO values.

It can be concluded from this study, as well as from other recent (29) and past studies (30), that the intrinsic growth rate of *Nitrosomonas* is not limited at DO concentrations above 1.0 mg/L, but that DO concentrations greater than 2.0 mg/L may be required in practice. When designing the aeration or oxygen addition component of a suspended growth nitrification system, it is recommended that a minimum DO level of 2.0 mg/L be specified at all times throughout the biological reactor to prevent peak load ammonia bleed-through. If significant, occasional transient conditions are anticipated, consideration should be given to providing standby DO capacity.

If mass-transport or diffusional resistances are an inherent characteristic of the nitrification reactor, as is the case with attached growth reactors, the DO level achievable in designing the oxygen addition component should be relatively high. Recent research work suggests that bulk fluid DO levels should be near 70 percent saturation. Lower levels may suggest mass-transfer limitations and limited ventilation (31). These considerations are discussed in more detail in Section 6.5.2.

3.3.4 pH and Alkalinity Effects

When the equation describing the complete nitrification process (Equation 3-8) is written in the context of the carbonic acid system, a substantial destruction of alkalinity is implied. It can be shown (4) that over a pH range of approximately 5 to 8 in an aqueous biological reactor, the equilibrium pH of the reactor will be dictated by the amount of alkalinity and CO_2 present in the system. Higher pH levels can be maintained at lower alkalinity levels in systems in which the stripping of CO_2 occurs in the biological reactor. Where the stripping of CO_2 does not occur, as is the case in enclosed systems, the alkalinity of the wastewater must be 10 times greater than the amount of ammonium nitrified in order to maintain a pH greater than 6.0 (32). Recall that the theoretical alkalinity destruction ratio is 7.1 mg (as $CaCO_3$)/mg of ammonium-nitrogen oxidized. The observed alkalinity destruction ratio has generally been equal to or less than the theoretical value in open systems using air as a source of oxygen (4).

Further information on the effect of particular aeration systems on the resulting reactor pH is provided in Section 6.4.10.4. The incorporation of a phosphorus removal capacity into nitrification systems through the addition of chemicals to the reactor(s) will also affect the reactor

alkalinity. Information on this consideration is presented in Section 6.4.10.2 and Table 2-3.

Reactor pH conditions have been found to have a significant effect on the rate of nitrification, as summarized in Figure 3-2. The degree of acclimation to the corresponding pH is also annotated on the figure. A wide range of optimum pH has been reported; an almost universal finding, however, is that as the pH moves to the acid range, the rate of ammonium oxidation declines. This tendency has been found to be true for both unacclimated and acclimated cultures, although acclimation, or selection of a different population of organisms with time, tends to moderate pH effects. In one study involving an attached growth reactor, nitrification declined by 50 percent at pH 6.0 after 1.5 d of acclimation, but no decline in nitrification performance was evident after acclimation for 10 d (32). In another study it was found that an abrupt change in reactor pH from 7.2 to 6.4 had no adverse effect on nitrification. However, when the pH was abruptly changed from 7.2 to 5.8, nitrification performance deteriorated markedly as effluent ammonium levels rose from approximately zero to 11 mg/L NH_4^+-N. A return to pH 7.2 caused rapid improvement, indicating that the lower pH was only inhibitory and not toxic (45).

For design purposes, it is sufficient to take into consideration that the nitrification rate may drop significantly as pH is lowered below the neutral range and that for performance stability it is best to maintain pH at 6.5–8.0. The effect of lower pH conditions, if they are anticipated, should not be ignored when sizing nitrification reactors, even though acclimation will attenuate the effect of pH on the nitrification rate.

3.3.5 Effect of Inhibitors

Nitrifying organisms are susceptible to a wide array of organic and inorganic inhibitors. As pointed out by Stover (46), nitrifiers can adapt to many inhibitory compounds when inhibitors are constantly present in the wastewater versus when slug discharges occur (e.g., from an accidental industrial discharge). Inhibition can occur through interference with the general metabolism of the cell or with the primary oxidative reactions. More important than distinguishing the mechanism of inhibition, however, is the need to establish a methodology for assessing the potential for, or occurrence of, nitrification inhibition in a biological system. Such procedures have been proposed by numerous researchers (46–48). More on design considerations that deal with the issue of nitrification inhibition is provided in Chapter 6 (Section 6.3.1).

Extensive reviews of the influence of selected inorganics and/or organics on nitrification inhibition have been prepared by Neufeld's group (49), Hockenbury and Grady (50), Pantea-Kiser's group (47), and Painter (51). While the data base on nitrification inhibition is extensive, Table 3-4 provides a list of several industrially significant or-
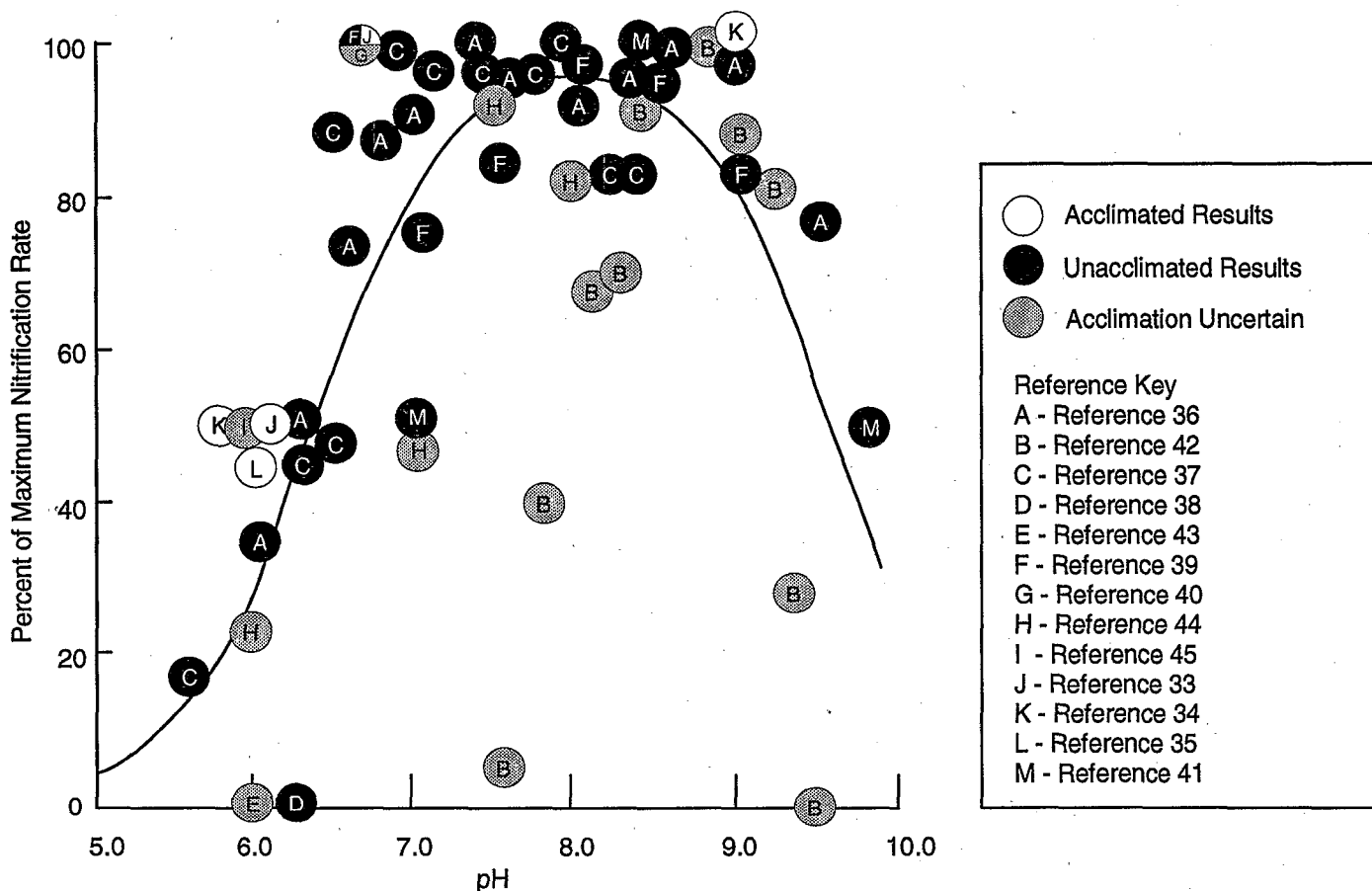
Figure 3-2. Effect of reactor pH conditions on rate of nitrification.

ganic chemicals found to cause some degree of nitrification inhibition. Care must be taken, however, when interpreting reported concentrations of inhibitory compounds, since acclimation can occur and effectively remove the inhibitory effect from a system; in a complete mixed system, the nitrifiers will normally see significantly lower concentrations than present in the influent, and suggested levels are often considerably higher than could occur in typical collection systems, particularly where pretreatment programs are in place. The reported data should be used as references for the relative effect of specific compounds.

Certain inorganics, including specific metals, are inhibitory to nitrifiers. Sawyer, on reviewing studies carried out in England, suggested that 10–20 mg/L of heavy metal

can be tolerated due to low ionic concentrations at pH values of 7.5–8.0. Inorganic compounds identified as potential inhibitors are listed in Table 3-5.

Nitrifying organisms are also sensitive to certain forms of nitrogen. Un-ionized ammonia (NH_3), or free ammonia (FA), and un-ionized nitrous acid (HNO_2), or free nitrous acid (FNA), are believed to be inhibitory to nitrifiers above certain concentrations. FA begins to inhibit *Nitrosomonas* at a concentration of 10–150 mg/L and *Nitrobacter* in the range of 0.1–1.0 mg/L (56). FNA begins to inhibit *Nitrosomonas* and *Nitrobacter* at concentrations of 0.22–2.8 mg/L. The FA and FNA concentrations are directly correlated to pH and temperature, and the concentration, respectively, of ammonia plus ammonium and nitrite plus

Table 3-4. Industrially Significant Organic Compounds Inhibiting Nitrification (Adapted from Reference 50)

Compound	Concentration of Compound Giving at Least 50 Percent Inhibition, mg/L
Acetone	2,000
Carbon disulfide	38
Chloroform	18
Ethanol	2,400
Phenol	5.6
Ethylenediamine	17
Hexamethylene diamine	85
Aniline	<1
Monoethanolamine	<200

Table 3-5. Metals and Inorganic Compounds Identified as Potential Nitrification Inhibitors

Compound	References
Zinc	1, 51
Free Cyanide	49
Perchlorate	1
Copper	1, 54
Mercury	1
Chromium	1, 53, 55
Nickel	1, 52, 54, 55
Silver	1
Cobalt	51
Thiocyanate	49
Sodium cyanide	52
Sodium azide	52
Hydrazine	52
Sodium cyanate	52
Potassium chromate	52
Cadmium	54
Arsenic (trivalent)	53
Fluoride	53
Lead	55

nitrous acid. FA and FNA are present in accordance with the following equilibrium reactions:

$$NH_4^+ + OH^- \leftrightarrow NH_3 + H_2O \qquad (3\text{-}15)$$

$$H^+ + NO_2^- \leftrightarrow HNO_2 \qquad (3\text{-}16)$$
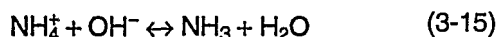
Threshold levels of ammonia plus ammonium-nitrogen, and nitrite plus nitrous acid-nitrogen at which nitrification inhibition may begin at a pH of 7.0 and a temperature of 20°C (68°F) are presented in Table 3-6 for illustrative purposes. (Values for other pH and temperature condi-

Table 3-6. Calculated Threshold Values of Ammonia Plus Ammonium-Nitrogen and Nitrite Plus Nitrous Acid-Nitrogen Where Nitrification Inhibition May Begin (from Reference 56)

Inhibitory FA or FNA Concentration, mg/L	Equivalent Ammonia plus Ammonium-N at pH 7.0 and 20°C, mg/L	Equivalent Nitrite plus Nitrous Acid-N at pH 7.0 and 20°C, mg/L
FA		
10 (*Nitrosomonas* Inhibition)	1,000	—
0.1(*Nitrobacter* Inhibition)	20	—
FNA		
0.22 (Nitrification Inhibition)	—	280

tions can be calculated [56].) The calculated values imply that it is unlikely that nitrification inhibition will occur as a result of the presence of ammonia plus ammonium and nitrite plus nitrous acid in the treatment of typical municipal wastewaters. However, sludge discharges into municipal systems of highly concentrated industrial wastes containing these forms of nitrogen can cause inhibition.

If anaerobic digestion is incorporated into a wastewater treatment plant and if untreated supernatant is returned to the process, a suitable reduction in the nitrification rate should be made. The growth rate of *Nitrosomonas* in a suspended growth reactor treating municipal wastewater can be inhibited by introduction of digester supernatant, according to Gujer (57) and others (58,59). Gujer's results indicate that the inclusion of digester supernatant recycle, to the extent that the ammonium-nitrogen concentration increases by 5 mg/L, can reduce the growth rate of *Nitrosomonas* by approximately 20 percent. The study assumed that the inhibiting compound was produced as a by-product of anaerobic degradation, versus any change in process conditions in the suspended growth reactor resulting from introduction of the digester supernatant.

3.3.6 Effect of Feed Organic Carbon to Nitrogen Ratio

The ratio of the feed biodegradable organic carbon, or CBOD, to the nitrogen available for nitrification in the wastewater (i.e., the C:N) is one of the critical factors affecting the design of nitrification systems. (A discussion of the C:N may also be found in Chapter 2 (Section 2.4.1.4).) Normally, for all nitrification systems, there is sufficient organic matter in the reactor feed to enable the growth of heterotrophic bacteria. Since the yield of heterotrophic bacteria is greater than the yield of the autotrophic nitrifying bacteria, there is a danger, when attempting to control the MLSS at a desired level, that the growth rate of the heterotrophic organisms will be

established at a value exceeding the maximum possible growth rate of the nitrifying organisms. Under such conditions the nitrifiers will be washed out of the system. Thus, in order to build and sustain a sufficient nitrifying population, the following condition must be satisfied, recognizing the relationship between growth rate and solids retention time:

$$\theta_c^d \geq \theta_c^m \qquad (3\text{-}17)$$

where:

θ_c^d = design solids retention time, d

θ_c^m = minimum solids retention time for nitrification at environmental conditions within the reactor, d

The net specific growth rate of the heterotrophic population can be expressed as:

$$\mu'_H = \frac{1}{\theta_c^d} = Y_H q_H - b_H \qquad (3\text{-}18)$$

where:

μ'_H = net specific growth rate of heterotrophs, d^{-1}

Y_H = true heterotrophic yield coefficient, g VSS grown/g of substrate CBOD (or COD) removed

q_H = rate of organic removal, g CBOD (or COD) removed/g of active VSS/d

b_H = decay coefficient, d^{-1}

The rate of organic removal is defined as:

$$q_H = \frac{S_0 - S_1}{X_1 t} \qquad (3\text{-}19)$$

where:

S_0 = feed total CBOD (or COD), mg/L
S_1 = effluent soluble CBOD (or COD), mg/L
t = reactor hydraulic retention time, d
X_1 = reactor active total VSS, mg/L

Methods have been proposed for estimating X_1 in Equation 3-19 by measuring the reactor total VSS (X) and correcting for inactive volatile solids as a result of endogenous decay (60,61), and accounting for the refractory VSS entering the reactor with the feed. The issue of predicting reactor total VSS has been dealt with recently in a more rigorous fashion in the activated sludge model development work of the IAWPRC, as referenced in Chapter 5. Predicting X can be simplified by using an observed or net biomass yield coefficient (Y_{NET}), as proposed in Chapter 2 (Figure 2-8). It is evident from Equations 3-17, 3-18, and 3-19 that once θ_c^d is selected, which is dictated by the anticipated growth rate of the nitrifiers, the required reactor hydraulic retention time and total VSS will depend on the feed biodegradable organic carbon concentration, since Y_H and b_H are assumed to be constant and S_1 will be at a minimal level in any reactor in which nitrification is occurring.

It is also evident from the preceding discussion that combining Equations 3-17, 3-18, and 3-19 not only illustrates how the feed organic carbon effects the nitrification reactor design, but effectively implies a procedure for design. The first step in the procedure is to define the design solids retention time. The use of this solids retention time approach for sizing suspended growth reactors is discussed further in Section 6.4.1.1.

An alternative to the solids retention time approach calls for determining a design ammonium oxidation or nitrification rate. Equation 3-11 can be expressed as follows if the nitrification rate is zero order ($K_N \ll N$) with respect to ammonium concentration:

$$q_N = \frac{\mu_N}{Y_N} = \hat{q}_N \qquad (3\text{-}20)$$

$$q_N = \frac{N_0 - N_1}{X_N t} \qquad (3\text{-}21)$$

where:

$N_0 - N_1$ = NH_4^+-N nitrified, mg/L

X_N = reactor active *Nitrosomonas* as VSS, mg/L

Equation 3-21 represents a simplified relationship for design, provided X_N can be determined. An estimate of X_N can be made with knowledge of the ratio of the feed biodegradable organic carbon and ammonium-nitrogen, and the active total VSS:

$$X_N = \frac{Y'_N (N_0 - N_1)}{Y'_N (N_0 - N_1) + Y_H (S_0 - S_1)} X_1 \qquad (3\text{-}22)$$

where:

Y'_N = organism yield coefficient for *Nitrosomonas* plus *Nitrobacter*, g VSS/g NH_4^+-N removed

Equation 3-22 assumes decay coefficients for nitrifiers and heterotrophs are negligible. It is clear from Equations 3-21 and 3-22 that wide variation in the maximum ammonium oxidation rate can be observed under similar environmental conditions, when the rate is calculated based on the measurement of total VSS. This variation is due to differences in the feed C:N. Methods are available to determine X_1 by measuring the reactor total VSS and correcting for the effect of organism decay and inert feed volatile solids. By combining Equations 3-13 and 3-20, the minimum solids retention time for nitrification, θ_c^m, can be related to the maximum ammonium oxidation rate:

$$\theta_c^m = \frac{1}{Y_N \hat{q}_N} \qquad (3\text{-}23)$$

This discussion suggests that using the nitrification rate approach can lead to errors in reactor sizing, unless the active solids basis to which the rate is normalized is explicitly understood. For this reason, the solids retention time approach to design has been favored.

3.3.7 Influence of Oxidative-Reductive Environments

Combining the processes of carbon oxidation, nitrification, denitrification and biological phosphorus removal in one or more reactors has become a favored technology application for control of nutrients from municipal wastewaters. In these systems, the nitrifiers are exposed to aerobic, anoxic (i.e., nitrate or nitrite present, but no DO), and anaerobic conditions. Anoxic or anaerobic zones are often incorporated into suspended growth systems designed for carbonaceous oxidation and nitrification to promote biological phosphorus removal and/or to control filamentous bulking. Recognizing the effect of DO on the nitrifiers raises concern about the design of nitrification systems in which the organisms are alternately exposed to conditions of low or zero DO. It has been reported that nitrifier activity is unaffected when exposed to anoxic conditions of up to 5 hr (62,63). The results from pilot plant studies at Burlington, Ontario, indicate that anaerobic conditions of up to 4 hr duration have no detrimental effect on nitrification rates (64). The incorporation of anoxic or anaerobic zones—referred to as selectors—to control filamentous bulking in full-scale suspended growth systems, had no reported effect on nitrification efficiency at selector hydraulic retention times of approximately 1.5 hr (65). Overall, available information leads to the reasonably safe conclusion that anoxic or anaerobic conditions lasting for hours have no tangible impact on nitrifier viability when acceptable DO concentrations are restored.

3.4 Attached Growth Kinetic Considerations

The kinetic description of fixed film system performance involves considerations beyond those presented in Section 3.3. Development of a particular biofilm reactor kinetic model requires applying the nitrification kinetic principles relevant for the nitrifying biomass to a reactor model that describes its hydrodynamics, mass-transfer characteristics, and any special features of the reactor. In contrast to suspended growth systems, transport processes are generally rate controlling in biofilm systems.

Mass-transport or diffusional resistances can or will influence the nitrification process in the following ways:

- They can negate the assumption that ammonium conversion to nitrite is the rate-limiting step in the nitrification process.

- They will increase the required reactor solids retention time.

- They will influence the value of DO in the bulk liquid at which the nitrification rate is limited.

These observations indicate that although the rate relationships presented in Section 3.3.1 and the qualitative relationships describing the effect of various environmental factors on nitrification are applicable to systems influenced by mass-transport or diffusional resistances, one must account for these influences. Diffusional resistances are of major concern in attached growth reactors in which growth occurs on or within a solid media. Since the design of this type of attached growth reactor is complicated by the need to consider the mass-transport processes and to define the film surface itself, it is primarily based on empirical results from pilot and full-scale systems. Nonetheless, reactor design relationships are consistent with biofilm models developed on the basis of stoichiometry, Fick's Law, and Monod Kinetics.

It has been found that the conditions in the immediate proximity of a microorganism in a biofilm are not the same as those measured in the bulk liquid. The concentration of substrates within the biofilm vary with depth and are significantly lower than in the bulk liquid, since they must be transported into and through the biofilm. Since this is true for DO and ammonium-nitrogen, lower concentrations within the biofilm can result in lower observed rates of nitrification than would be predicted based on liquid ammonia concentrations and the amount of attached biomass.

Figure 3-3 is a schematic representation of the film theory model normally used to represent a biofilm. The model considers both external and internal transport limitations. The incomplete mixing of the bulk liquid with the liquid phase immediately adjacent to the biofilm surface indicates that external mass-transfer resistance is an important consideration. External transport limitations occur when substrate must diffuse through a stagnant liquid film at the attached growth surface. For diffusion to occur, a concentration gradient must exist. Flux of material through this layer is generally modeled as follows:

$$J = AD(\Delta S/\Delta L) \qquad (3\text{-}24)$$

where:

J = the flux, mass/time
A = the biofilm surface area, length^2
D = the diffusion coefficient of the component of interest, length^2/time
ΔS = the difference in substrate concentration between the bulk liquid and the liquid film at the biofilm surface, mass/length^3
ΔL = the thickness of the stagnant film, length

As indicated in Equation 3-24, the concentration of substrate at the surface will be lower than the concentration in the bulk liquid. Moreover, after reaching the biofilm surface, the substrate must diffuse through the biofilm to reach the microorganisms within the attached growth matrix. This step will further reduce the substrate concentration within the biofilm since it requires a concentration gradient.

Some of the observations that can be made based on the biofilm model are of interest when considering am-
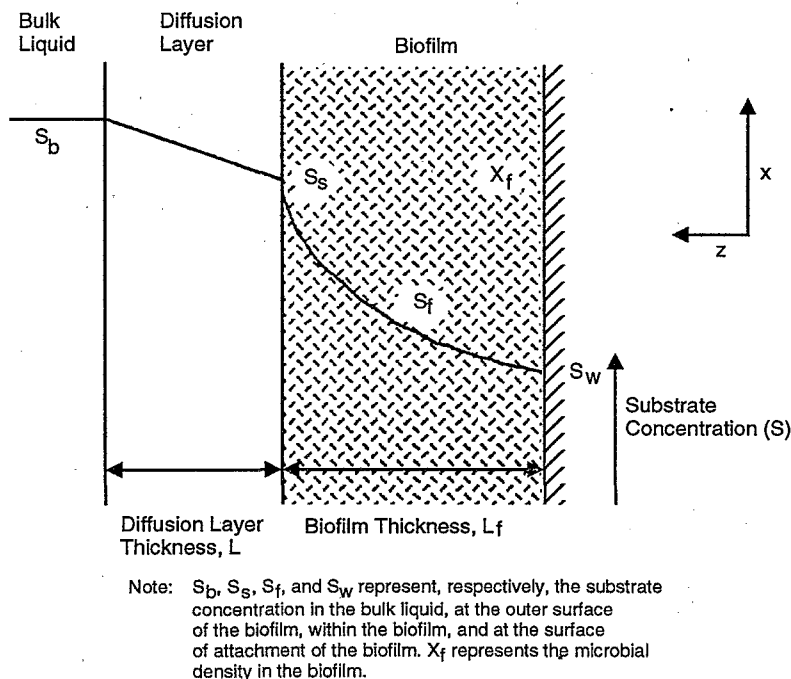
Figure 3-3. Schematic of conceptual biofilm model (adapted from Reference 70).

Note: S_b, S_s, S_f, and S_w represent, respectively, the substrate concentration in the bulk liquid, at the outer surface of the biofilm, within the biofilm, and at the surface of attachment of the biofilm. X_f represents the microbial density in the biofilm.

monium removal rates in attached growth systems. The biofilm model shows that the ammonium oxidation rate in attached growth systems should not be decreased as drastically under adverse environmental conditions as in suspended growth systems (66). This finding is consistent with the observation made elsewhere that attached growth systems have an advantage over suspended growth systems in withstanding lower temperatures (4,67,68). The biofilm model also shows that the bulk liquid DO concentration must be 2.7 times the ammonium-nitrogen concentration to prevent oxygen transfer from limiting nitrification rates in attached growth systems (66). Operational procedures suggested to overcome this limitation include dilution of the ammonium concentration through recirculation and increasing oxygen transfer through the use of high-purity oxygen (66).

Numerous models have been developed to describe nitrification and other biological processes in biofilm reactors; the performance modeling of these systems is addressed in detail elsewhere (69). The models generally consider microbial kinetics and diffusion through the biofilm simultaneously. The solution of general biofilm models, which consist of a set of partial differential equations, is generally infeasible even with the best numeric techniques. As such, biofilm models are normally formulated to answer specific questions of interest. Even then, however, such models are relatively complex and their solutions often obtainable only through the use of time-consuming numerical techniques (70). Although advances in the application of such models have resulted from the development of pseudo-analytical solutions, de-

signs based on the use of such models are still more complex than designs for reactors in which mass-transport resistances can be neglected, as is the case for most suspended growth reactors. No simple design criteria with a theoretical basis such as solids retention time are available for the design of attached growth nitrification processes, although attempts have been made to relate the solids retention time in a trickling filter to the media surface loading rate (71,72). Further information concerning modeling approaches is presented in Section 6.5.1. In the absence of simple, widely accepted design criteria that are theoretically based, the relationships presented for design of the various alternative attached growth reactor configurations discussed in Chapter 6 are less theoretically precise than those developed for suspended growth reactors.

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National Technical Information Service
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Chapter 4
Process Chemistry and Kinetics of Biological Denitrification

4.1 Introduction

This chapter reviews the fundamentals of the chemistry, biochemistry, and kinetics of denitrification, focusing on the treatment process. The discussion seeks to provide an understanding of the underlying principles affecting the performance, design, and operation of denitrification processes. Subsequent chapters deal specifically with the design and operation of these processes.

Biological denitrification involves the microbial reduction of nitrate to nitrite, and ultimately nitrite to nitrogen gas. Nitrate and nitrite replace oxygen for microbial respiration in this reaction; as such, denitrification is commonly thought to occur only in the absence of molecular oxygen. The conditions suitable for denitrification—oxygen is absent but nitrate is present—are commonly referred to as anoxic.

Since nitrogen gas is relatively biologically inert, denitrification converts nitrogen from a potentially objectionable form (nitrate) to a form that has no significant effect on the environment (nitrogen gas). As discussed in Chapter 1, nitrate in water can be objectionable if nutrient enrichment is a concern and/or if the water is intended to be potable. Denitrification in wastewater treatment applications may also provide process benefits in certain situations, including the development of alkalinity, the reduction of oxygen demand, and production of an activated sludge with better settling characteristics.

4.2 Fundamentals of Denitrification

4.2.1 Microbiology

Unlike nitrification, a relatively broad range of bacteria can accomplish denitrification. Denitrifiers are ubiquitous in most natural environments, including municipal wastewaters and sludges (1,2). Many of the microorganisms in municipal activated sludge systems are denitrifiers, even in systems that are not specifically designed for denitrifying. The presence of the organisms is due in part to the fact that they are facultative: they can use either oxygen or nitrate as their terminal electron acceptor. Denitrifiers can proliferate in aerobic systems because of their ability to use oxygen and efficiently oxidize organic matter (2).

The ubiquity of denitrifiers minimizes the need to create special environmental conditions for their survival, as must be done for nitrifiers.

4.2.2 Metabolism and Biochemical Pathways

In the process of denitrification, nitrate and nitrite act as electron acceptors in the respiratory electron transport chain in the same manner as oxygen. This transport chain is the fundamental mechanism by which cells generate energy. The process involves transferring electrons from a reduced electron donor (e.g., an organic substrate) to an oxidized electron acceptor (e.g., oxygen, nitrate, nitrite, or sulfate). Nitrate or nitrite may serve as a substitute for oxygen in this chain with only small modifications to the metabolic system (i.e., the enzymes) of the bacteria. By using nitrate or nitrite in place of oxygen in the electron transport chain, however, slightly less energy is generated. Similarly, more energy is generated using nitrate than sulfate.

Control systems exist within individual bacteria and natural microbial populations that ensure the most efficient form of energy generation is utilized. Thus, if oxygen is present, it will be used preferentially over nitrate, and if oxygen is not present, nitrate will be used preferentially over sulfate. Since the bacteria that reduce sulfate (sulfate reducers) cannot compete effectively with nitrate reducers for the available organic matter, sulfate reduction to sulfide and resulting odor production are not likely to occur in a treatment system that is anoxic (i.e., where nitrate is present). Also, significant sulfate reduction is not likely to occur in a system that may be void of oxygen and nitrate for short periods of time (e.g., the few hours in the anaerobic zone of a biological phosphorus removal activated sludge system), since the sulfate reducers will not have adequate time to proliferate in the numbers required to carry out significant sulfate reduction. Moreover, sulfate reducers may be poisoned in the aerobic zones of such systems.

The control mechanism in denitrifiers that allows them to switch from oxygen to nitrate occurs at two levels. The first is the synthesis of the enzymes required for denitrification. In pure cultures, oxygen has been found to repress the synthesis of these enzymes. Between 2 and

3 hr is typically required for pure cultures to synthesize the enzymes when the cells are switched from an aerobic to an anoxic environment. Research on activated sludge, however, suggests that the denitrifying enzymes may be present even in systems that do not have anoxic conditions (3), suggesting that a period for synthesis of the denitrifying enzymes is not required prior to initiation of denitrification.

Control of denitrification also occurs at the level of enzyme activity. Oxygen will inhibit the activity of the denitrifying enzymes. The oxygen concentration at which denitrification stops has been reported to be 0.2 mg/L in pure cultures (4). In activated sludge systems the reported values are 0.3 to 1.5 mg/L (5), possibly because of the diffusion limitations of microbial floc (4).

Research also suggests some species of bacteria can concurrently use nitrite and oxygen as terminal electron acceptors (6). Studies by Kugelman et al. (7) suggest that concurrent oxygen and nitrite utilization in activated sludge systems operated at high DO levels is possible. This phenomenon was observed in activated sludges grown in systems with anaerobic/aerobic conditions, such as those encountered in biological phosphorus removal systems. Similar results were not observed in sludges grown in a purely aerobic nitrifying system. At present, data are insufficient to indicate whether concurrent oxygen utilization and nitrite reduction is a general phenomenon with anaerobic/aerobic systems and how this phenomenon might be exploited.

Total nitrogen losses of up to 30 percent are very often observed in aerobic nitrifying activated sludge systems. These losses may be due to concurrent oxygen and nitrate/nitrite utilization, denitrification occurring inside the activated sludge floc under anoxic conditions, or denitrification occurring in clarifiers and other unaerated zones. Due to the unpredictable nature of these losses, they are not generally relied upon in the design of systems for nitrate removal.

The list of organic compounds that can serve as organic substrates (i.e., as carbon and electron donors) for denitrification is especially long. Almost any compound that is degraded with oxygen as the electron acceptor will also serve as an electron donor with nitrate. Some aromatic compounds that do not serve as electron donors for nitrate exist (e.g., benzene), possibly because of the need for oxygen in the enzymatic cleavage of the aromatic ring. The organic compounds that could be used for the denitrification of wastewaters include:

- Organics present in municipal and industrial wastewaters
- Methanol
- Ethanol
- Acetic acid

- Waste organic material

While organics in wastewater and methanol are the two most commonly used electron donors, electron donor selection will depend primarily on cost and local availability.

4.2.2.1 Theoretical Stoichiometric Equations

Theoretical stoichiometric equations can be used to predict the mass of electron donor (i.e., carbon substrate) and acceptor (i.e., oxygen, nitrate, or nitrite) consumed, and the mass of cells produced during any biological process. Since denitrification involves the transfer of electrons from donor to acceptor, half-reactions can be used to develop the stoichiometric equations. Half-reaction equations for oxygen, nitrate, and nitrite as electron acceptors are as follows:

$$\frac{1}{4} O_2 + H^+ + e^- \rightarrow \frac{1}{2} H_2O \qquad (4\text{-}1)$$

$$\frac{1}{5} NO_3^- + \frac{6}{5} H^+ + e^- \rightarrow \frac{1}{10} N_2 + \frac{3}{5} H_2O \qquad (4\text{-}2)$$

$$\frac{1}{3} NO_2^- + \frac{4}{3} H^+ + e^- \rightarrow \frac{2}{3} H_2O + \frac{1}{6} N_2 \qquad (4\text{-}3)$$
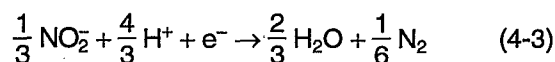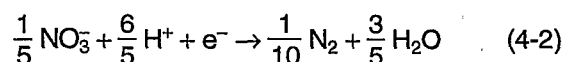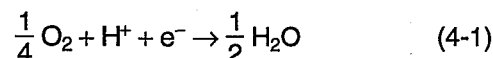
A comparison of equations 4-1 and 4-2 indicates that 8 g of oxygen (1/4 x 32 g O_2/mole) is equivalent to 2.86 (1/5 x 14 g-N/mole) of nitrate nitrogen. Hence, the reduction of 1 g of nitrate nitrogen is equivalent to the reduction of 2.86 g of oxygen. Using this stoichiometric equivalence, the reduction in oxygen demand that is possible when denitrification is incorporated into a single-sludge activated sludge system (i.e., raw wastewater is used as the carbon and energy source) can be estimated. Since 4.6 g of oxygen are required to oxidize ammonia nitrogen to nitrate nitrogen (Equation 3-3), and 2.86 g of oxygen equivalents are recovered in the denitrification of the nitrate nitrogen, it is theoretically possible to reduce the net energy expended in providing oxygen for nitrification by up to 63 percent by using the raw wastewater for denitrification. Since this factor could provide significant energy savings, it should be taken into account in evaluating flow scheme alternatives.

Reduction of the electron acceptors (oxygen, nitrate, or nitrite) requires an electron donor, which can either be the organic substrate in the raw wastewater or a substrate added to the source. The most commonly used external carbon source is methanol when denitrification is accomplished as a separate stage. The half-reaction equation for methanol as the electron donor is:

$$\frac{1}{6} CH_3OH + \frac{1}{6} H_2O \rightarrow \frac{1}{6} CO_2 + H^+ + e^- \qquad (4\text{-}4)$$
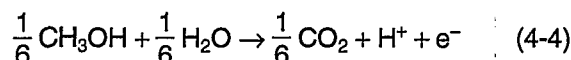
Half-reactions also can be written for a variety of other organic compounds serving as electron donors (8,9).

The reactions in Equations 4-2 and 4-4 can be combined as follows:

$$NO_3^- + \frac{5}{6} CH_3OH \rightarrow \frac{5}{6} CO_2 + \frac{1}{2} N_2 + \frac{7}{6} H_2O + OH^- \quad (4\text{-}5)$$
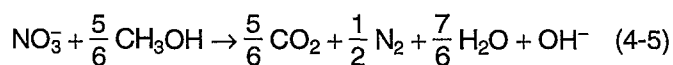
According to this relationship, 1.9 g of methanol (2.86 g when expressed as COD) are required per g of nitrate-nitrogen reduced.

In natural water systems, the carbonic acid buffering system is the dominant inorganic component of the water. Consequently, the hydroxide (OH⁻) produced during denitrification in natural waters will react with carbonic acid (carbon dioxide) to produce bicarbonate ions (i.e., bicarbonate alkalinity). The following equation takes this conversion into account (10):

$$NO_3^- + \frac{5}{6} CH_3OH + \frac{1}{6} H_2CO_3 \rightarrow \frac{1}{2} N_2 + \frac{4}{3} H_2O + HCO_3^-$$

$$(4\text{-}6)$$
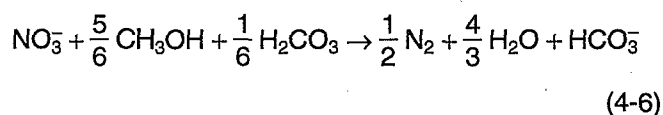
In addition to supplying electrons to produce energy, the organic substrate also provides the carbon for creating new cell mass in heterotrophic microorganisms; consequently, more electron donor will be required to reduce a given amount of nitrate than is presented in Equation 4-5. The distribution of the organic substrate between incorporation into new biomass versus that utilized for the reduction of nitrate will be a function of the organic compounds making up the substrate, the microbial population present, and the operating conditions of the system. Thus, this distribution must be determined experimentally.

The experiments of McCarty et al. (10) provide the basis to formulate stoichiometric equations for methanol consumption incorporating both the half-reactions above and the observed biomass yields. These equations are as follows (11):

$$NO_3^- + 1.08 \, CH_3OH + 0.24 \, H_2CO_3 \rightarrow$$
$$0.056 \, C_5H_7NO_2 + 0.47 \, N_2 + 1.68 \, H_2O + HCO_3^- \quad (4\text{-}7)$$

$$NO_2^- + 0.67 \, CH_3OH + 0.53 \, H_2CO_3 \rightarrow$$
$$0.04 \, C_5H_7NO_2 + 0.48 \, N_2 + 1.23 \, H_2O + HCO_3^- \quad (4\text{-}8)$$

$$O_2 + 0.93 \, CH_3OH + 0.056 \, NO_3^- \rightarrow$$
$$0.056 \, C_5H_7NO_2 + 1.04 \, H_2O + 0.59 \, H_2CO_3 +$$
$$0.056 \, HCO_3^- \quad (4\text{-}9)$$
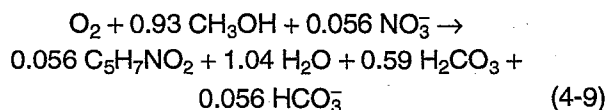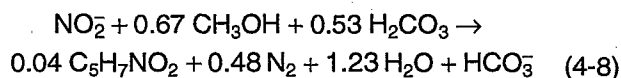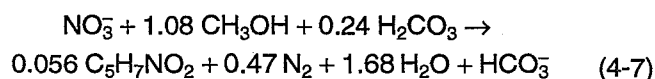
In converting to mass, Equation 4-7 suggests 2.47 g of methanol are required to reduce 1 g of nitrate-nitrogen. Experimental ratios of methanol to nitrate range from 2.5 to 3.0 g methanol/g nitrate nitrogen. Methanol may also be required to reduce any nitrite or oxygen present. The following equation has been used to describe the overall methanol requirement (10):

$$M = 2.47 \, (NO_3^-\text{-}N) + 1.53 \, (NO_2^-\text{-}N) + 0.87 \, DO \quad (4\text{-}10)$$

where:

M = methanol required, mg/L
NO_3^--N = nitrate nitrogen removed, mg/L
NO_2^--N = nitrite nitrogen removed, mg/L
DO = DO removed, mg/L

Departure of methanol requirements from Equation 4-10 is most likely due to variations in sludge yields among experimental systems. An expression defining the required COD to nitrogen ratio has been developed (12) for any organic substrate:

$$\frac{COD}{N} = \frac{2.86}{1 - 1.134 \, Y_{S_{NET}}} \quad (4\text{-}11)$$

where:

$Y_{S_{NET}}$ = biomass net yield based on COD, g VSS/g COD removed

This equation was developed assuming that the COD of the VSS produced is 1.42 g COD/g VSS, and that the biomass produced is 10 percent nitrogen.

A second method that can be used to obtain a rough estimate of the organic substrate requirements is to consider the oxygen demand exerted in a given system by an organic substrate and convert it to nitrate equivalence based on a factor of 2.86 mg oxygen/mg nitrate nitrogen. For example, if it is known from experience for a given organic substrate and treatment system that the oxygen demand exerted is 200 mg/L, then 70 mg/L of nitrate nitrogen could theoretically be reduced.

4.2.2.2 Alkalinity and pH Relationships

As mentioned above, bicarbonate alkalinity is produced and carbonic acid concentrations are reduced during denitrification. The theoretical stoichiometry of the bicarbonate alkalinity production is 3.57 mg alkalinity as $CaCO_3$ produced per mg of nitrate nitrogen reduced to nitrogen gas. The preponderance of investigative evidence strongly suggests that the actual denitrification alkalinity production is essentially equivalent to stoichiometric predictions.

Since the alkalinity concentration is increased and the carbonic acid concentration is reduced, the tendency of denitrification is to partially reverse the effects of nitrification and raise the pH of the biological reaction. For waters low in alkalinity, the recovery of alkalinity through denitrification in a single-sludge system can be a significant benefit. The recovery of alkalinity and the potential reduction in energy requirements attributable to denitrification may in some situations make denitrification attractive even if total nitrogen limits are not required.

103

4.3 Kinetics of Denitrification

This section presents a partial development of equations that can be used to describe the rate of denitrification. The discussion is intended to show the relationship between denitrification and the most prominent rate-limiting factors. The material serves as the basis of more detailed, process-specific kinetic equations development in Chapters 5, 7, and 8.

4.3.1 Biomass Growth and Nitrate Utilization

The kinetics of denitrification can be described using equations that take the same form as those for other microbial reactions, such as nitrification or organic matter removal. For example, first-order, zero-order, and Monod-type kinetics have all been used to describe the rate of denitrification (Section 3.3.1). However, the development of one set of kinetic expressions to cover the general topic of denitrification design is not possible given the varied nature of the configurations used to accomplish denitrification and the different organic substrates used. The expression used will often be a function of the type of reactor utilized, the organic substrate employed, and the needs of the user.

Monod-type expressions are developed first in this discussion in order to illustrate the general concepts of denitrification kinetics. These expressions are often directly applicable to separate denitrification reactors with supplemental substrate addition (e.g., methanol). For single-sludge denitrification systems, the expressions are more difficult to apply directly because of factors such as the heterogeneity of the substrate and the biomass. Expressions that may be used for single-sludge denitrification systems are briefly discussed in this chapter and again in Chapter 8. The IAWPRC model described in Chapter 5, however, is based on Monod-type expressions.

A Monod-type expression can be used to relate the growth rate of denitrifying microorganisms to the concentration of nitrate:

$$\mu_D = \hat{\mu}_D \left(\frac{D}{K_D + D} \right) \qquad (4\text{-}12)$$

where:

μ_D = specific denitrifier growth rate, d^{-1}
$\hat{\mu}_D$ = maximum specific denitrifier growth rate, d^{-1}
D = concentration of nitrate nitrogen, mg/L
K_D = half-saturation coefficient, mg/L

Such an expression will transform into a zero-order expression ($\mu_D = \hat{\mu}_D$) if K_D is considerably less than D, and a first-order expression ($\mu = (\frac{\hat{\mu}_D}{K_D}) \times D$) if D is significantly less than K_D.

As discussed in Section 3.3.1, the net specific growth rate of microorganisms in a system is the inverse of the solids retention time (SRT):

$$\frac{1}{\theta_c} = \mu' \qquad (4\text{-}13)$$

where:

θ_c = solids retention time, d
μ' = net specific growth rate, d^{-1}

Since the growth rate of denitrifiers is generally very similar to aerobic heterotrophic organisms (i.e., much greater than nitrifiers), the minimum solids retention time required to prevent washout of denitrifiers from a reactor will be much shorter than for nitrifiers.

Nitrate removal rates can be related to organism growth rate by using the organism yield as a conversion factor:

$$q_D = \frac{\mu_D}{Y_D} = \hat{q}_D \left(\frac{D}{K_D + D} \right) \qquad (4\text{-}14)$$

where:

q_D = nitrate removal rate, g NO_3^--N/g VSS/d
Y_D = denitrifier true yield coefficient, g biomass (VSS) grown/g NO_3^--N removed
$\hat{q}_D$ = maximum nitrate removal rate, g NO_3^--N/g VSS/d

The solids retention time can be related to nitrate removal rates in a similar fashion:

$$\frac{1}{\theta_c} = Y_D q_D - b_d \qquad (4\text{-}15)$$

where:

b_d = denitrifier decay coefficient, d^{-1}

The concentration of organic substrate will also affect the rate of nitrate removal. Thus, the rate of nitrate removal can also be expressed as a function of organic substrate concentration:

$$q_D = \hat{q}_D \left(\frac{S}{K_S + S} \right) \qquad (4\text{-}16)$$

where:

S = concentration of organic substrate, mg/L
K_s = half-saturation constant, mg/L

Since both nitrate and electron donor concentrations may limit the rate of denitrification, a more accurate expression for the rate is a double (or multiple) Monod-type expression (9,11):

$$q_D = \hat{q}_D \left(\frac{S}{K_S + S} \right) \left(\frac{D}{K_D + D} \right) \qquad (4\text{-}17)$$

The values of the half-saturation constant for nitrate, K_D, are reported to be very low: from about 0.1 to 0.2 mg

NO$_3^-$-N/L (1,13,14). It can be seen from examination of Equation 4-17 with these values for K_D that nitrate concentrations greater than 1 to 2 mg NO$_3^-$-N/L have almost no effect on denitrification rates, provided diffusion limitations are not a factor. Consequently, nitrate concentrations will not affect the rate of nitrate removal unless very low nitrate effluent concentrations must be achieved or diffusion limitations exist such as may be encountered in attached growth processes.

Values for K_s will depend on the organic substrate, but are also generally low. Values of K_s for methanol of 0.1 mg/L (11) to 6.0 mg/L (7), and as high as 72 mg/L (15), have been reported. The low K_s value using methanol implies that to achieve 90 percent of the maximum denitrification rate, only about 1 mg/L of methanol need be in the effluent; that is, significant excesses of methanol above stoichiometric requirements need not be in the effluent to approach the maximum denitrification rate. This factor can be of significance in some situations since excess methanol in the effluent may require removal in a downstream aerobic process. Although very limited data are available in the literature, it is likely that the values of K_s with municipal wastewater as the organic substrate (expressed as BOD) are likely to be greater than those with methanol. Consequently, nitrate removal rates are likely to be a function of the organic substrate concentration in a system using wastewater as the organic substrate, as will be discussed below.

The impact of oxygen on denitrification can also be added to denitrification rate expressions as follows (16):

$$q_D = \hat{q}_D \left(\frac{S}{K_S + S} \right) \left(\frac{D}{K_D + D} \right) \left(\frac{K_o}{K_o + S_o} \right) \qquad (4\text{-}18)$$

where:

K_o = half-saturation constant for oxygen, mg/L
S_o = DO concentration, mg/L

The term $K_o/(K_o+S_o)$ can act as a switching function, turning denitrification on and off. A value for K_o of 0.1 mg/L has been suggested for denitrification in the IAWPRC model (see Chapter 5).

The kinetics of denitrification can also be expressed in terms of the rate of organic substrate removal, since the nitrate removal is tied directly to the organic removal. In many instances use of the substrate removal rate may be preferred, since more information is available on kinetic coefficients for substrate removal. The equations will be identical in form to those above, but the rates of nitrate removal will be replaced by the rates of substrate removal. For example, the rate of substrate removal can be expressed as:

$$q_s = \hat{q}_s \left(\frac{S}{K_s + S} \right) \qquad (4\text{-}19)$$

where:

q_s = substrate removal rate, g COD or BOD$_5$/g VSS/d
$\hat{q}_s$ = maximum substrate removal rate, g COD or BOD$_5$/g VSS/d

The value of K_s will remain the same as that used for the nitrate removal rate (see Equation 4-16).

The rate of nitrate removal can be related to the rate of substrate removal with the same type of expression commonly used to relate oxygen consumption to organic substrate utilization (12):

$$q_D = (1 - 1.42\,Y_s) \frac{q_s}{2.86} + \frac{1.42}{2.86} b_d \qquad (4\text{-}20)$$

where:

Y_s = biomass true yield, g biomass (VSS) grown/g COD removed

In this equation, the first term is used to describe the fraction of the substrate (expressed as COD) and nitrate that are used for biomass synthesis, and the second term describes the fraction of nitrate that is used for endogenous respiration. The value 1.42 is an assumed value for the COD of the VSS produced.

Table 4-1 summarizes some of the kinetic coefficients that have been determined for suspended growth systems with methanol as the carbon source. In most cases only net (or observed) yields are reported or can be calculated from the available data. The classical relationship (Table 2-13) between the true yield (due to biomass growth) and the net yield is:

$$Y_{D_{NET}} = \frac{Y_D q_d - b_d}{q_D} = \frac{Y_D}{(1 + \theta_c b_d)} \qquad (4\text{-}21)$$

where:

$Y_{D_{NET}}$ = denitrifier net (observed) yield, g VSS/g NO$_3^-$-N removed

A value for b_d of 0.04 d^{-1} was used in some cases to derive calculated Y_D values for those cases where none was reported. There is significant variation in the data represented in Table 4-1, due to differences in test conditions and procedures. Care should be taken when using these data without consulting the specific reference.

Further, care must be taken when using these data or any others reported in the literature to determine the basis for the specific rates. Data may be reported on the basis of active denitrifying biomass or total volatile solids. Those listed in Table 4-1, for instance, are for the most part based on total volatile solids; however, the rate expressions provided generally in this chapter are based on active denitrifying biomass. In actual systems the VSS will be made up of active autotrophic bacteria (nitrifiers), nondenitrifying active heterotrophic bacteria, denitrifying

Table 4-1. Values for Denitrification Yield and Decay Coefficients for Various Investigations Using Methanol (from Reference 18)

q_D, d^{-1}	$Y_{D_{NET}}$	Y_D	b_d, d^{-1}	Temp., °C	$\hat{q}_s$	Y_s	K_s
Variable	Variable	0.57	0.05	10			
Variable	Variable	0.63	0.04	20			
Variable	Variable	0.67	0.02	30			
0.12 to 0.32	0.55 to 1.4	—	—	20			
0.16 to 0.9	0.57 to 0.73	—	—	20			
Variable	0.53	—	—	20			
—[a]	0.58	0.77[b]	0.04[c]	10 to 20			
0.131 to 0.347	0.542 to 0.703	0.84	0.04	20			
0.25	0.49	0.65[b]	0.04[c]	16 to 18			
—[d]	0.7 to 1.4	0.83 to 1.67	0.04[c]				
0.30	0.61	0.65[c]	0.19[b]	19 to 20			
0.12 to 0.24	—	—	—	20			
	—	—	0.04	10	3.1	0.17	12.6[f]
	—	—		20	10.3	0.18	9.1
0.1 to 0.45[e]	—	—	—	16 to 23			
0.3 to 0.56	—	—	—	16 to 18			

[a] q not given, but θ_c = 8.0
[b] Calculated
[c] Assumed
[d] θ_c = 3 to 6 days
[e] With a 1-hour aerobic stabilization step
[f] Based on total COD (methanol plus any nonbiodegradable material)

q_D = Nitrate removal rate
$Y_{D_{NET}}$ = Net yield based on nitrate = g VSS/g NO_3-N removed
Y_D = True yield based on nitrate = g VSS/g NO_3-N removed
b_d = Endogenous decay coefficient
$\hat{q}_s$ = Maximum specific COD removal rate = g COD/g VSS/d
Y_s = True yield based on COD = g VSS/g COD
K_s = COD half-saturation constant (based on total COD) = mg/L COD

active heterotrophic bacteria, inert organics in the influent, and inert products of biomass decay. The IAWPRC model described in Chapter 5 takes these different fractions into account. For more simplified design approaches, total VSS may be used provided the data were obtained from studies on a system with similar wastewater characteristics, so that the various fractions that make up the VSS are the same. Alternatively, rates specific to active biomass may be used if the fraction of the total active VSS can be estimated. Table 2-13 provides the IAWPRC equation that can be used to make this estimation.

While a number of methods can be used for sizing a denitrifying reactor with supplemental substrate addition, the method chosen will be a function of readily available kinetic data and the type of reactor used. For a completely mixed, suspended growth reactor, the rate of nitrate or substrate utilization (i.e., that which can be estimated from Monod-type kinetics) can be used in conjunction with a mass balance equation to calculate a design reactor volume. An alternative approach (but one that is based on the same fundamental equations) is to calculate a design solids retention time from the rate of nitrate or substrate utilization (Equation 4-15). Similar procedures

can be used from that point as would be used for a nitrifying activated sludge system (See Chapters 3 and 6). For a suspended growth system, the solids retention time should be checked to verify that it is greater than the minimum (1.0 to 2.5 d) required to develop a flocculent biomass.

The kinetic equations developed above may be readily applied to systems where a single, readily biodegradable electron donor is used (e.g., methanol). The application of the equations is somewhat more complicated in systems that have a complex electron donor source (e.g., municipal and many industrial wastewaters). Often only part of the organic matter in these wastewaters is readily biodegradable with the remaining fractions being slowly biodegradable, or nonbiodegradable. The slowly biodegradable fraction may be particulate or soluble.

Also, the kinetic expressions above may be applied to complex electron donor sources by assuming that the growth of denitrifiers (and heterotrophs in general) only occurs at the expense of the readily biodegradable fraction and that the slowly biodegradable fraction must be converted ("hydrolyzed") to a readily biodegradable form

prior to its utilization (16). Separate kinetic expressions can be used to express the rate at which the slowly biodegradable organics are converted to a readily biodegradable form. In addition, growth of denitrifiers will also occur at the expense of biodegradable organic matter produced in the reactor from the decay of other cells—a phenomenon termed endogenous respiration. As previously indicated, kinetic equations may also be used to express the rate at which microorganisms decay. All of these processes are incorporated into the single-sludge activated sludge model prepared by IAWPRC (16), as discussed in Chapter 5.

Simplified approaches based on empirical observation can be used for designing the denitrification process in single-sludge systems with complex substrates. It has been observed that the process of nitrate removal exhibits distinct rates that are a function of the availability of the substrate (i.e., the location of the anoxic zone in the process) (7,21). Ekama (21) noted two phases of denitrification in a plug flow, first anoxic zone and a third phase in a second anoxic zone. The first phase was fairly short (up to 10 minutes) and may have been due to the consumption of the readily degradable organic fraction. Burdick et al. (5) observed only one rate in the first anoxic zone of Bardenpho facilities. The rate was correlated to the food/biomass (F/M) loading on the anoxic zone:

$$SDNR_1 = 0.03 \left(\frac{F}{M_1} \right) + 0.029 \qquad (4\text{-}22)$$

where:

$SDNR_1$ = specific denitrification rate in the first anoxic zone, g NO_3^--N/g VSS/d

F/M_1 = F/M loading ratio on the first anoxic zone, g BOD/g mixed liquor VSS in the first anoxic zone/d

Specific rates of denitrification in the first anoxic zone typically range from 0.04 to 0.15 g NO_3^--N/g VSS/d (22). This range and Equation 4-22 provide only rough estimates of the denitrification rate that will be observed, since the rate will be influenced by a number of factors including the mixed liquor recirculation rate, the influent wastewater strength, and the denitrifying active fraction of the volatile solids. The expressions used in the IAWPRC model (see Chapter 5) are designed to take some of these factors into account to give a more accurate estimate of the actual denitrification rate.

Denitrification in second anoxic zones generally is driven by the endogenous decay of the biomass, so that the rates will be much lower than in the first anoxic zone. Specific rates of denitrification in the second anoxic zone are typically 20-50 percent of the rate in the first zone (11,22). The rate of endogenous decay will be a function of the system and the SRT. Two equations that have been developed to describe the rate in the second anoxic zone are provided below (5,12):

$$SDNR_2 = 0.12 \times \theta_c^{-0.706} \qquad (4\text{-}23)$$

$$SDNR_2 = \frac{0.175 \, A_n}{\hat{Y} \, \theta_c} \qquad (4\text{-}24)$$

where:

$SDNR_2$ = specific denitrification rate in the second anoxic zone, g NO_3^--N/g VSS/d

$\hat{Y}$ = net TSS production across the entire activated sludge system, including inert solids and biomass, g TSS/g BOD_5 removed

A_n = net amount of oxygen required across the entire activated sludge system/TBOD removed, g O_2/g BOD

4.3.2 Temperature Effects

As with any microbial activity, nitrate removal rates can be affected significantly by temperature. Data from the literature are summarized in Figure 4-1, where they have been normalized with respect to rate at 20°C (68°F). These data suggest that temperature exerts a larger effect below about 20°C (68°F) than above.

The impact of temperature on biological systems is often described by a Arrenhius-type function:

$$q_{D,T} = q_{D,20} \, \theta^{(T-20)} \qquad (4\text{-}25)$$

where:

$q_{D,T}$ = denitrification rate at temperature T(°C), mg NO_3^--N/g VSS/d

$q_{D,20}$ = denitrification rate at 20°C, g NO_3^--N/g VSS/d

θ = simplified Arrenhius temperature-dependent constant

Although such a function is useful for modeling denitrification, it is limited to a certain temperature range, and the value of theta (θ) is site specific. Values of theta reported in the literature are summarized in Table 4-2.

Table 4-2. Temperature Correction Coefficients for Modeling Denitrification (Endogenous Rate)

θ	Reference
1.08	23
1.09	24
1.20	25
1.08	25
1.03	25
1.08	26
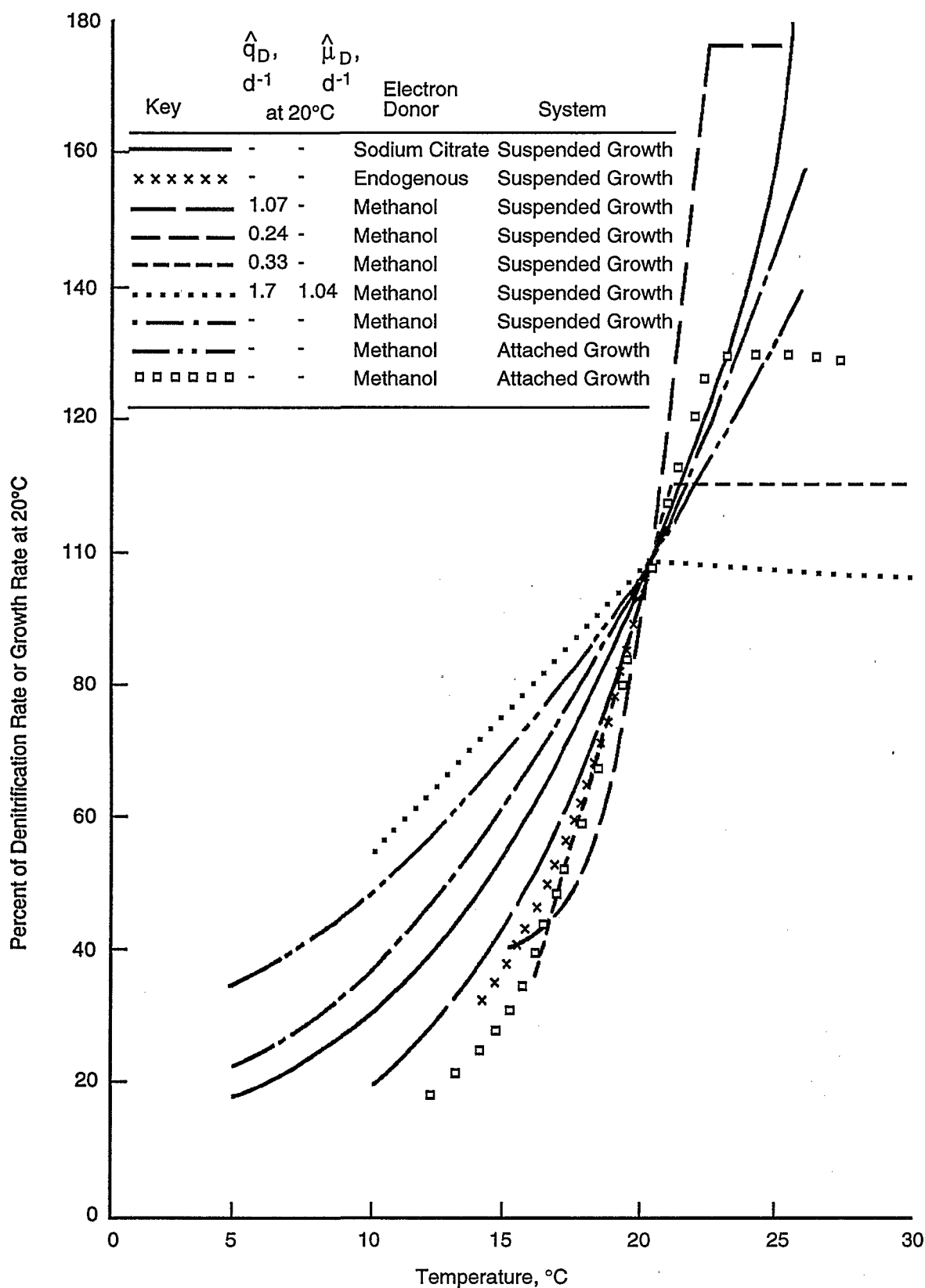
Figure 4-1. Effect of temperature on denitrification rates (from Reference 11).

4.3.3 pH and Alkalinity Effects

The response of denitrification and aerobic respiration rates to pH variations should be similar. In general, denitrification will be much less sensitive to pH than nitrification. Representative observations of the effect of pH on denitrification rates are shown in Figure 4-2. These data suggest that denitrification rates are depressed below pH 6.0 and above pH 8.0. Since denitrification will produce alkalinity, it may increase the pH if high concentrations of nitrate are to be removed.

4.3.4 Effects of Inhibitors

Denitrifiers are much less sensitive to inhibitory compounds than are nitrifiers. In general, inhibitors would be expected to have a similar degree of impact on denitrification and heterotrophic aerobic respiration. Consequently, commonly applied concentrations that result in inhibition (e.g., those published by EPA for activated sludge and trickling filters [27]) can be used for denitrification. The ability of a biomass to acclimate to higher levels of inhibitory compounds should be taken into account when reviewing these values. Much higher concentrations may be tolerated by acclimated cultures. Specific literature should be reviewed or pilot tests conducted to determine actual inhibitory levels.

4.3.5 Effect of Diffusional Limitations

Diffusional limitations will affect the design of fixed film reactors for denitrification as they will for fixed film reactors for nitrification (see Section 3.4). In general, the equations presented above are applicable to fixed film systems only if they are coupled with equations that describe diffusional limitations. Models have been developed that take such limitations into account (see Chapter 5). However, the design of many attached growth reactors is quite often based on empirical results from pilot and full-scale systems. Rates of denitrification in these empirical approaches are often based on media surface area or media volume.

For reactors that provide very turbulent conditions, such as fluidized beds, the rate of mass transport may be so high that diffusion may not limit the rate of reaction. The design of such reactors may be based on the same equations as those used for suspended growth reactors, provided the biomass in the reactor can be estimated.
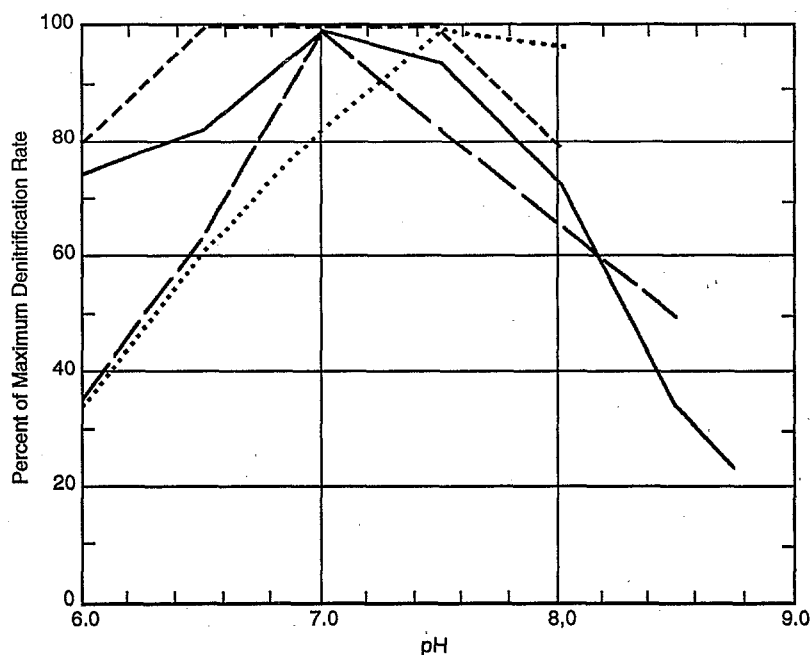
Figure 4-2. Effect of pH on denitrification rates (from Reference 11).

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When an NTIS number is cited in a reference, that document is available from:

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Chapter 5
Mathematical Modeling of Nitrification and Denitrification

5.1 Introduction

Mathematical modeling is a technique that is increasingly being used to analyze problems of significance to environmental engineering. This analytical approach can be used as an adjunct to more traditional tools (e.g., those described in Chapters 6, 7, and 8) to refine and optimize proposed facility designs. It can also be used as an operational tool to optimize the performance and/or capacity of an existing facility. Moreover, appropriate mathematical models can be used to analyze a wider range of options, for both facility design and operation, than would be possible by other means, resulting in more informed decision-making. Analysis of facility design options can lead to better decisions concerning facility size and operational flexibility needs. For example, mathematical modeling can be used to optimize the oxygen transfer system configuration for a single-sludge nitrification/denitrification system. Similarly, modeling can be used to evaluate a variety of operational strategies for an existing system, allowing the selection of the most attractive strategies for full-scale evaluation.

This chapter discusses available approaches for modeling nitrification/denitrification systems. The discussion is organized so that modeling approaches and the different types of models are discussed first. Next, requirements to develop a mathematical model for a particular application are discussed, followed by a review of uses of mathematical models. Finally, mathematical models currently available for use in the design and evaluation of nitrification and denitrification systems are presented, and an example illustrates the use of a mathematical model to optimize the design of a single-sludge nitrification/denitrification system.

The discussion in this chapter independently addresses the topic of mathematical modeling. Although a firm understanding of nitrogen removal technology is necessary to utilize fully the material presented here, thorough review of the other chapters of this manual is not necessary for the knowledgeable practitioner to benefit from the discussion. Users of this manual may elect to proceed directly to Chapters 6, 7, and 8 if system design is of more immediate interest.

5.2 Modeling Approaches

A wide variety of models is used to analyze problems in environmental engineering. For the purposes of this discussion, however, the models are divided into two categories: conceptual models and mathematical models.

5.2.1 Conceptual Models

Conceptual models form the basis for most of the design and operational decisions made in environmental engineering. A conceptual model represents an understanding of the cause-and-effect relationships at work in a given system. Figure 5-1, for example, illustrates a conceptual
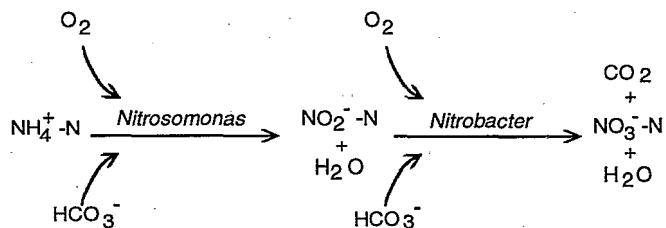
Figure 5-1. Conceptual nitrification model.

model for the nitrification process. Ammonium nitrogen (NH_4^+-N) is first oxidized to nitrite nitrogen (NO_2^--N) through the action of autotrophic bacteria, most often of the genus *Nitrosomonas*. Molecular oxygen (O_2) and bicarbonate alkalinity are consumed as a result of this reaction; reaction by-products include water (H_2O) and additional bacteria. NO_2^--N generated in the first step is subsequently oxidized to nitrate nitrogen (NO_3^--N) through the action of a second type of autotrophic bacteria of the genus *Nitrobacter*; co-reactants and by-products for this reaction are similar to those for the first reaction.

Although not quantitative in nature, this model can be used to understand many of the factors that affect the nitrification process in wastewater treatment systems. For example, it is clear from Figure 5-1 that nitrification will not occur unless sufficient quantities of the two types of bacteria—*Nitrosomonas* and *Nitrobacter*—are present. Concentrations will be sufficient only if the environmental conditions allow these bacteria to proliferate. Factors

111

such as temperature, pH, DO, and specific growth rate affect the concentration of the nitrifying bacteria and, based on empirical observations and correlations, conditions that encourage growth of sufficient numbers of these bacteria in various wastewater treatment systems can be identified. Using this information, it is possible to design and operate wastewater treatment systems to accomplish nitrification (the conversion of NH_4^+-N to NO_3^--N).

Based on this example, the characteristics of a conceptual model can be identified:

- *System components.* Conceptual models identify the components of a system. In Figure 5-1, for example, the system components include NH_4^+-N, O_2, bicarbonate alkalinity, *Nitrosomonas,* NO_2^--N, H_2O, *Nitrobacter,* and NO_3^--N. Conceptual models, however, present one representation of how a system might actually function.

- *Cause-and-effect relationships.* Conceptual models describe the cause-and-effect relationships between system components. For example, consider the interaction of the two bacterial species required to accomplish nitrification. Figure 5-1 indicates that NH_4^+-N is first oxidized to NO_2^--N by the action of *Nitrosomonas.* Subsequently, the generated NO_2^--N is oxidized to NO_3^--N by the action of *Nitrobacter.* This relationship indicates that both species must be present in the appropriate amounts to obtain complete nitrification. Other relationships, such as the effect of aeration on the loss of CO_2 and the subsequent increase in reactor pH, can also be similarly described. To the extent possible, the proposed interrelationships typically represent actual physical, chemical, and/or biological phenomena.

- *Interactions can be described qualitatively and/or quantitatively.* The relationships illustrated in Figure 5-1 are qualitative. However, because the stoichiometry of the reactions presented is known, they could also be presented in quantitative form. Qualitative information, such as an increase in X leads to an increase in Y under certain conditions but to a decrease in Y under alternate conditions, can also be presented. Interrelationships may also be presented as generalized correlations (e.g., effluent NH_4^+-N concentrations are correlated with suspended growth system solids residence time as a function of temperature). In this latter instance, physical data are presented graphically.

Conceptual models, as described above, can and have been used routinely to design nitrogen removal systems. Designs produced based on the use of these models can be highly efficient and cost effective. However, as discussed below, design approaches based on mathematical modeling procedures can enhance designs developed using more conventional analytical tools.

5.2.2 Mathematical Models

Mathematical models have many features in common with conceptual models. The primary difference is that in a mathematical model the interrelationships between system components must be described quantitatively, while in a conceptual model they may be described either qualitatively or quantitatively. Consequently, mathematical models are in one sense a subset of conceptual models. However, the restriction that mathematical models can use only quantitative relationships can lead to significant practical differences between conceptual and mathematical models.

Because of the complexity of the physical, chemical, and biological systems being modeled, simplifications are frequently introduced into mathematical models to reduce the mathematical complexity. These simplifications take several forms. In some instances, the impacts of several interactions are included in a single factor. For example, the single half-saturation constant in the Monod-type substrate concentration-reaction rate relationship (the type often incorporated into many mathematical models) is often used to represent the impact of more than one variable. In other instances, approximations are used as mathematical conveniences to reduce the computations required to produce a mathematical solution to a problem. Thus, mathematical models often represent simplifications of the current understanding of the physical problem being considered. While some simplifications are introduced to reduce the computational complexity of a proposed problem, others are introduced because of the difficulty in measuring the stoichiometric and kinetic coefficients required by the model.

Mathematical models can be used as a tool to test alternative hypotheses concerning the interrelationships between system components. Alternative models can be developed based on contrasting hypotheses, and system responses can then be predicted under defined conditions. Through comparison of detailed model predictions with the results of actual measurements for bench, pilot, and/or full-scale systems, the utility of the proposed models can be analyzed and the most useful one selected. Although close correspondence between model predictions and actual responses does not prove the subject hypothesis, it does demonstrate that the hypothesis is useful. This approach can be used as a tool to test alternative hypotheses and to select the one that best represents a particular application. Used in this fashion, mathematical modeling can lead to enhanced insight concerning both the interactions between process variables affecting nitrogen removal systems and the relative importance of various process variables in determining overall process performance.

Although increased computer power, provided by the new generation of computers, is making the use of complex simulation models possible, it is important to recognize

the value of simplicity. As a general rule, it is best to use the simplest model that provides an adequate description of the system to be analyzed. Simplicity often leads to clarity in thinking and evaluation, and to the avoidance of errors caused by a lack of understanding of the fundamental features that determine system response. Thus, while the computer provides the ability to simultaneously consider many factors, this capability must be balanced against the value of gaining a basic understanding of a system. As is typically the case, such issues can be resolved best using the direct experience of the practitioner.

Many types of mathematical models can be applied to engineering applications. For instance, statistical models are used to analyze existing data sets as a way to identify and quantify interrelationships between process variables. Semi-empirical models are used to correlate process variables and facilitate engineering calculations. In contrast, mechanistic models are based on a specific interpretation of a phenomenon occurring within a system being modeled. In the following discussion, emphasis is placed on mechanistic models because they are based on phenomena that determine system response, while also offering the best potential for providing a realistic and unbiased representation of the system being considered. They also offer the best opportunity to accurately predict system response over a broad range of operating conditions.

5.3 Model Development

Since mechanistic mathematical models are developed to achieve specific objectives, different models will be used to achieve different purposes. Model development proceeds most efficiently if a logical approach is used. One approach consists of three steps: model selection, model calibration, and model verification. These are illustrated in Figure 5-2 and are discussed in this section.

5.3.1 Model Selection

The first step in model development is model selection. In this step, the specific form of the model to be used for a particular application is selected. Factors to be considered include the existing and possible level of knowledge concerning the application (i.e., the data available), the decisions to be made based on the modeling results, and the available understanding of system mechanics.

The available data affect model selection since they determine the detail possible for the modeling exercise. For example, if it is not possible to obtain a·detailed characterization of the subject wastewater, it makes little sense to use a sophisticated model capable of providing detailed predictions about the impacts of wastewater characteristics on system response. Predictions based on assumed characteristics would be of little value since they will not, in all likelihood, reflect actual operating condi-
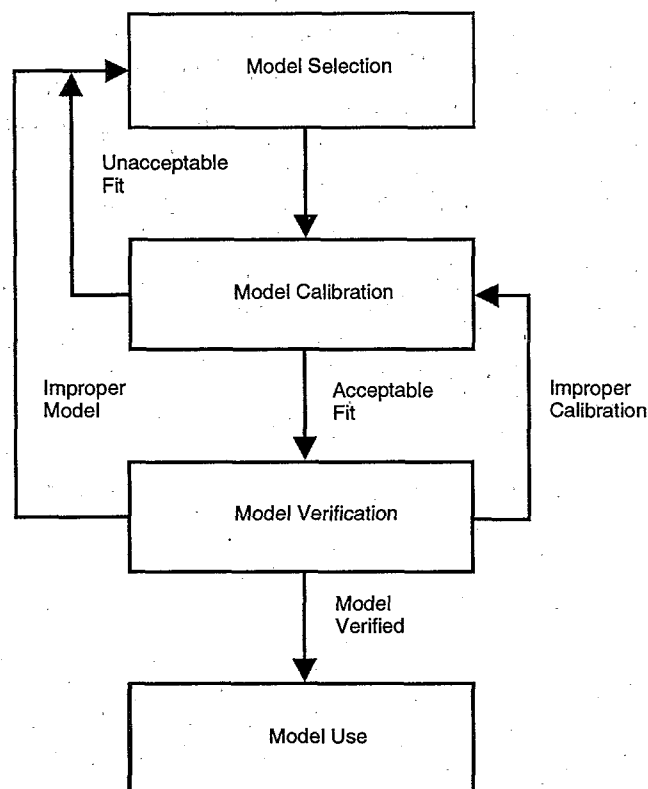
Figure 5-2. Stepwise approach to model development.

tions. In such a case, use of a generalized model would be more appropriate.

The use to which the model is to be put will also determine model·structure. A relatively simple model can be used to size an aeration basin that would allow a complete mix activated sludge system to accomplish single-stage nitrification. In contrast, a more sophisticated model is required to size a single-sludge nitrification/denitrification system, particularly one using staged reactors for the anoxic and aerobic zones. Even more sophistication is required if spatial and temporal variations in process oxygen requirements are to be estimated.

A final consideration in model selection is the level of current knowledge concerning the underlying mechanics of the system being simulated. Mathematical models are one means to represent the current understanding of a specified system. As such, the predictive capabilities are no better than the fundamental knowledge that is available on that system. In other words, mathematical modeling builds on existing knowledge and, although it can be used to test alternative hypotheses, it cannot be used by itself to expand existing knowledge concerning a particular system.

Many different models could be applied to a particular engineering problem. The knowledge and judgement of the process modeler, coupled with experience with alter-

native process models, form the basis for the initial selection of the model type for a particular application.

5.3.2 Model Calibration

Once a model has been selected, it must be calibrated. The objective of calibration is to select appropriate values for the required kinetic and stoichiometric coefficients. Selection can be based on several factors. In some cases, the coefficients are fundamental in nature and can be derived from theoretical considerations. In others, significant experience has been acquired with use of the particular model in similar applications, and that experience can be used as the basis for the selection of model coefficients. In still other cases, the model is calibrated by adjusting model coefficients to match the results of bench, pilot, and/or full-scale facilities that simulate the proposed treatment system. The data used for this approach can vary widely, from a few simple bench scale measurements to the results from sophisticated pilot plants, prototypes, and/or full-scale facilities. When experimental results are used to calibrate a proposed model, however, the experiments must have been designed to produce a range of responses that is sufficiently broad to allow for accurate determination of model coefficients. Mathematical models are most accurate when used to interpolate within the existing data base; extrapolations must be approached with caution. In short, a rational procedure must be used to select the numerical values of model coefficients if realistic results are to be obtained.

Model calibration also involves evaluation of the adequacy of model structure. If the model is unable to properly predict trends in system response and/or if model coefficients cannot be adjusted sufficiently to obtain an acceptable calibration, it may be necessary to adjust the structure of the model. Again, model responses must be tested over a sufficiently broad range of conditions to ensure its suitability for the intended application. Sensitivity analysis is one tool that can be used to assess model structure. In such an analysis, model imputs are varied and the resulting impact on model results is determined. Based on the results, the significance of various model components can be assessed. If the model is found to be lacking, model selection must be reconsidered. This so-called feedback loop is considered in Figure 5-2. Once a new model is selected, it must again be calibrated.

5.3.3 Model Verification

The final step in model development is model verification. In this step, predictions by the calibrated model are compared to measured results from a prototype system to identify similarities and differences between predicted and measured results. Verification can be considered valid only if the data set used for verification is different from the data set used for calibration. (Verification using an independent set of data is necessary because a properly calibrated model will always show acceptable results when compared to the data used for calibration. If acceptable results are not obtained using calibration data, then the model is inappropriate or it has not been properly calibrated.) The model is considered to be validated when model predictions agree with measured values from an independent data set within acceptable tolerances. Discrepancies between predicted and measured values indicate improper model selection and/or calibration, dictating that one or both of these steps must be repeated. Once a revised and/or recalibrated model is identified, it should be verified with yet another set of independent data. In some instances, additional independent data will not be available and the process modeler must judge whether a model appropriate for its intended use has been produced.

As with model calibration, the data set used for model verification must be sufficiently broad to test the calibrated model over the entire range of the potential application.

5.4 Model Applications

As discussed above, the level of complexity appropriate for a particular model is dependent on the intended use for the model. This section discusses potential uses for nitrification and denitrification process models.

5.4.1 New Facilities

One use of a mathematical model is to design new nitrogen control facilities. In such uses, a great deal of judgement must be applied in the selection of an appropriate model. If there is a lack of information on the characteristics of the wastewater to be treated, then a relatively simple modeling approach may be most appropriate. On the other hand, if significant data are available based on experience with the particular system and/or the type of wastewater to be treated, then a more sophisticated modeling approach may be utilized. In many cases, model verification will be difficult to carry out with new facilities because of a lack of an appropriate data base. A significant pilot plant program may need to be conducted if a sophisticated model is necessary to optimize the design of a new facility.

Extreme care must be exercised in designing a pilot plant program that is intended to produce the data necessary to calibrate a sophisticated process model. Factors of particular concern include:

- *Pilot plant configuration.* The configuration of the pilot plant should fully simulate the type of facility to be evaluated. At the least, the hydraulic flow pattern, oxygen transfer characteristics, and other performance characteristics should be similar to the proposed facility.

- *Pilot plant operating conditions.* Operating conditions for the pilot plant should simulate anticipated full-scale

conditions. Factors such as diurnal variations in process loadings, seasonal variations in wastewater characteristics (including temperature), and the impacts of in-process recycle streams (especially those from solids handling systems) should be considered.

- *Experimental design for model calibration.* A broad range of operating conditions should be explored in the pilot plant program to develop a sufficiently broad data base to allow accurate estimation of model parameters. Mathematical models are best used to interpolate within a data set; extrapolation beyond the collected data base should be used cautiously, if at all.

Sensitivity analysis and statistical design are important concepts in regard to the experimental design of the pilot plant program. A sensitivity analysis is accomplished by varying the numerical value of the parameters in the process model and determining the resulting impact on the predicted response. A parameter with high sensitivity is one for which a small variation results in a large variation in the response predicted by the model. A parameter with low sensitivity is one that may be varied over a relatively wide range while producing only a relatively small variation in the predicted response. Although accurate estimates are required for the sensitive parameters if the model is to provide accurate predictions, less accuracy is needed in the estimates for the relatively insensitive parameters. Consequently, the pilot plant program should be designed specifically to provide accurate estimates of the sensitive parameters.

Experimental design is the selection of target operating conditions for the pilot plant, and variables to be measured that will allow accurate estimation of the high-sensitivity parameters. The process model itself can be used as a tool to design the experimental program. The model can be run over a wide range of potential operating conditions and parameter values, and the results can be used to select target operating conditions that will provide the most accurate numerical estimate of the model parameters. It can also be used to identify system responses that are the most useful in model calibration. For example, process oxygen requirement values are often more sensitive to suspended growth model parameters than is process effluent quality. Consequently, calibration of the model based on measured process oxygen requirements can lead to more accurate calibration of the model than measurement of process effluent quality. The process model can be used to facilitate development of the most efficient experimental program. It can also be used to identify variables that are the most useful in accurately determining the numerical values of model parameters.

Significantly, estimates of model parameters based on pilot plant data are "coupled," not absolute. While error in the estimation of one parameter is often compensated for by a corresponding error in the estimated value of another parameter, the magnitude of such errors is un-

known. As a result, measured values of model parameters must be considered relative to each other, not as independent values. Thus, the absolute value of a model parameter is seldom known and the relative value, in combination with the relative value of all the other model parameters, must suffice to produce model predictions. This consideration is particularly important when estimates of model parameters obtained from one pilot plant study are compared to or used as inputs for other studies.

5.4.2 Existing Facility Upgrading

Upgrading of existing wastewater treatment facilities often represents better opportunities to apply sophisticated mechanistic mathematical models to facility design. In such cases, information on wastewater characteristics is available to facilitate model calibration. However, if the proposed upgrade represents a significant increase in the level of treatment, the performance characteristics of the existing system may not closely match those of the proposed system. While the proposed model may be calibrated to the existing system, the actual operating conditions for the existing system may vary considerably from those associated with the proposed design. Consequently, the opportunities for model verification using data from the existing system may be limited. (The discussion of experimental design for model calibration, presented in Section 5.3.2, is relevant in such cases.) One advantage in modeling an existing facility for an upgrade, in comparison to modeling a new facility, is that the existing wastewater and the biomass may be used for some limited bench scale verification runs. Batch tests can be designed using the subject model, as discussed above. If warranted, the availability of the existing wastewater can allow operation of a pilot plant utilizing the proposed process.

5.4.3 Existing Facility Optimization or Expansion

Optimization or expansion of an existing facility represents the best opportunity for use of mathematical modeling approaches. The proposed model can be calibrated and the results verified using data from the existing full-scale system. The highest degree of calibration and verification is obtained in these instances since the existing full-scale operation is the system to be studied. Modeling can be used as a tool to identify opportunities to optimize the existing system; both operational and physical modifications can be evaluated. The most desirable modifications, based on the modeling results, can then be evaluated in the full-scale system. If results comparable to those predicted by the model are obtained, the model can be further used for facility optimization and/or to identify facility expansion opportunities. If results do not compare favorably, the model can be modified accordingly and the process repeated. In any event, the capabilities and limitations of the model can be carefully defined.

A calibrated mathematical model of an existing treatment system can also be used on an ongoing basis as a tool

to optimize facility operations. Alternative operational strategies can be evaluated using the mathematical model, and the most promising approaches can then be implemented on a full-scale basis. The successive interaction between use of the model to predict system responses and monitoring of actual system responses can lead to development of an invaluable tool for maximizing the performance and/or capacity of an existing system.

5.5 Available Models

A wide variety of models are available to characterize full-scale nitrification and denitrification systems. Since models for suspended growth systems are the most fully developed and tested, they are discussed in great detail in this section. Also, however, ongoing research directed at the development of improved models for fixed film systems is summarized. It is hoped that efforts in this area will produce useful models that can be applied to the design and operation of fixed film nitrogen control systems.

5.5.1 Suspended Growth Models

While a wide variety of models has been developed to characterize suspended growth nitrification and denitrification systems, the most complete model was developed by a task group sponsored by the International Association on Water Pollution Research and Control (IAWPRC) (1,2). This model, referred to as Activated Sludge Model No. 1, was developed specifically for use in evaluating long sludge age (i.e., nitrifying) activated sludge systems, including those incorporating single-sludge denitrification features. It incorporates features found useful in previous models; consequently, detailed description of this model allows coverage of the relevant features of other, less sophisticated models. A computerized version of the IAWPRC model is available in the public domain (3). Key features of the model, which include wastewater characterization, biomass fractions, model stoichiometry and kinetics, and model presentation, are discussed below.

The notation system used to present the model was developed by another IAWPRC task group. This system is presented in the literature (4).

5.5.1.1 Wastewater Characterization

Key aspects of Activated Sludge Model No. 1 include the use of chemical oxygen demand (COD) to characterize the organic matter in wastewater, rather than BOD_5, and the division of the organic matter into various fractions. COD is used rather than BOD_5 to allow straightforward calculation of process oxygen requirements. If both COD and BOD_5 data are available for a particular wastewater, development of design COD values is easily accomplished. If only BOD_5 data are available, it must be converted to biodegradable COD values for input to the model. Division of the organic matter into various fractions allows use of a structured modeling approach, which is

necessary to accurately predict spatial and temporal variations in carbonaceous process oxygen requirements.

Figure 5-3 presents the division of organic matter utilized by Activated Sludge Model No. 1. The organic matter is first divided into biodegradable and nonbiodegradable fractions. Nonbiodegradable organic matter is further divided into soluble (S_I) and particulate (X_I) fractions according to the different processing of the two materials in the activated sludge system. Soluble, nonbiodegradable materials simply pass through the system and appear in the effluent at a concentration equal to the influent concentration. In contrast, particulate, nonbiodegradable materials are enmeshed into the activated sludge mixed liquor and accumulate as sludge, which is captured in the secondary clarifier and returned to the biological reactor. The concentration factor for a suspended growth system is the ratio of the solids residence time (SRT) to the hydraulic residence time (HRT) of the system and often represents an approximate 20-fold concentration factor.

Biodegradable organic matter is also divided into two fractions: readily biodegradable (S_S) and slowly biodegradable (X_S). The model assumes that slowly biode-
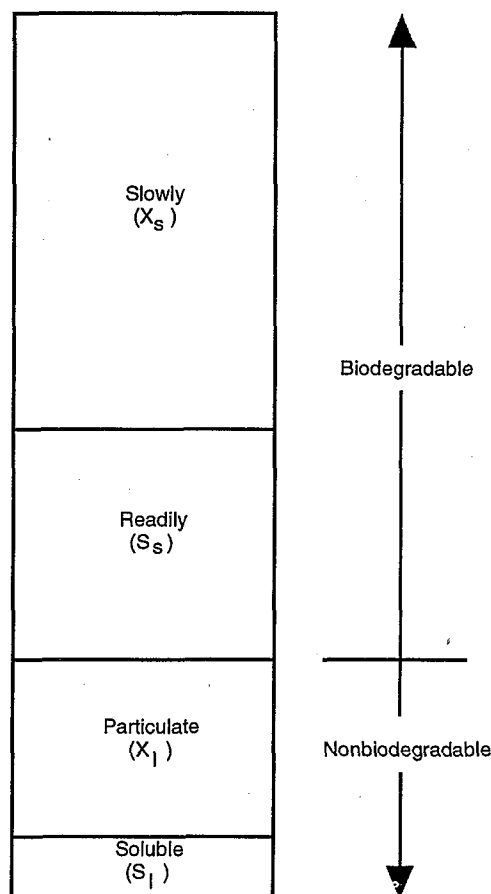
Figure 5-3. Division of organic matter in Activated Sludge Model No. 1.

gradable organic matter is first hydrolyzed into readily biodegradable organic matter that is subsequently consumed by heterotrophic bacteria. Note that these fractions *do not* in general correspond to the division of biodegradable organic matter into soluble and particulate fractions (see Reference 5 for a more detailed discussion of this topic). Keep in mind that 1) the readily degradable organic matter will typically be less than the soluble organic matter, as traditionally defined by standard practice; and 2) biological methods are generally superior to physical/chemical methods for characterizing the readily biodegradable organic matter in a wastewater stream.

In general, the biodegradable organic matter content (readily plus slowly biodegradable components) and the measured BOD_5 of a wastewater are related. In fact, they represent different approaches to characterizing the same components. In the absence of site-specific data, it may be assumed that the biodegradable organic matter content is equal to the ultimate BOD (BOD_{ult}).

Nitrogen components are expressed as nitrogen (N) and are divided into inorganic (ammonium, S_{NH}, and nitrate plus nitrite, S_{NO}) and organic (soluble biodegradable, S_{ND}, and particulate biodegradable, X_{ND}) fractions. Nitrogen fractions are converted into their oxygen equivalents when they are involved in oxidation/reduction reactions. This conversion occurs internal to the model. The stoichiometric factors are 4.57 mg O_2/mg NH_4^+-N oxidized and 2.86 mg O_2/mg NO_3^--N reduced. Nondegradable organic nitrogen (both soluble and particulate materials) is treated in the same manner as nondegradable organic matter. DO (S_O) and alkalinity (S_{ALK}, molar units) are also included in the model.

5.5.1.2 Biomass Fractions

Three biomass fractions are included in the model: active heterotrophic bacteria ($X_{B,H}$), active autotrophic bacteria ($X_{B,A}$), and the inert products of biomass decay (X_P). Heterotrophic bacteria include those organisms that are both capable and incapable of denitrification. Correction factors are used to account for 1) growth of heterotrophic bacteria (η_g) and 2) hydrolysis of slowly biodegradable organic matter (η_h) under anoxic conditions. These factors account both for the fraction of heterotrophic bacteria that are capable of denitrification and for any difference in activity from aerobic to anoxic conditions. The factors are empirical in nature and, as such, their numerical values should be confirmed for a particular application. Growth of active heterotrophic bacteria occurs when DO and/or nitrate nitrogen is present; no growth occurs when both DO and nitrate nitrogen are absent. Autotrophic bacteria include both *Nitrosomonas* and *Nitrobacter* which grow only when DO is present.

The model uses an unconventional approach to biomass decay. Active biomass (both heterotrophic and autotrophic) is assumed to decay into slowly biodegradable or-

ganic matter (X_S) and inert product (X_P) in a process that does not require oxygen or nitrate. Oxygen (or nitrate) utilization occurs only when the slowly biodegradable organic matter is subsequently hydrolyzed and then metabolized by heterotrophic bacteria. This has two impacts: first, the decay rate must be much higher than values traditionally used in order to produce oxygen consumption and net biomass production rates consistent with those observed in practice; second, biomass decay is assumed to occur under all conditions, but the resulting hydrolysis of slowly degradable organic matter (X_S) occurs only under aerobic or anoxic conditions. Consequently, slowly biodegradable organic matter will accumulate under anaerobic conditions.

Table 5-1 summarizes the components of the model, along with their descriptions.

Table 5-1. Activated Sludge Model No. 1: System Components

Component Number	Component Symbol	Definition
1	S_I	Soluble inert organic matter, $M(COD)L^{-3}$
2	S_S	Readily biodegradable substrate, $M(COD)L^{-3}$
3	X_I	Particulate inert organic matter, $M(COD)L^{-3}$
4	X_S	Slowly biodegradable substrate, $M(COD)L^{-3}$
5	$X_{B,H}$	Active heterotrophic biomass, $M(COD)L^{-3}$
6	$X_{B,A}$	Active autotrophic biomass, $M(COD)L^{-3}$
7	X_P	Products from biomass decay, $M(COD)L^{-3}$
8	S_O	DO, $M(COD)L^{-3}$
9	S_{NO}	Nitrate and nitrite nitrogen, $M(NL)L^{-3}$
10	S_{NH}	Ammonia nitrogen, $M(N)L^{-3}$
11	S_{ND}	Soluble biodegradable organic nitrogen, $M(N)L^{-3}$
12	X_{ND}	Particulate biodegradable organic nitrogen, $M(N)L^{-3}$
13	S_{ALK}	Alkalinity, Molar units

M = mass
L = length

5.5.1.3 Stoichiometry and Kinetics

Table 5-2 summarizes the stoichiometric and kinetic parameters used in the model, while Table 5-3 summarizes the kinetic expressions. Aerobic growth of heterotrophic bacteria includes Monod expressions for readily biode-

Table 5-2. Activated Sludge Model No. 1: Kinetic and Stoichiometric Parameters

Kinetic event	Symbols
Heterotrophic growth and decay	$\hat{\mu}_H$, K_S, $K_{O,H}$, K_{NO}, b_H
Autotrophic growth and decay	$\hat{\mu}_A$, K_{NH}, $K_{O,A}$, b_A
Correction factor for anoxic growth of heterotrophs	η_g
Ammonification	k_a
Hydrolysis	k_h, K_X
Correction factor for anoxic hydrolysis	η_h

Stoichiometric coefficient	
Heterotrophic yield	Y_H
Autotrophic yield	Y_A
Fraction of biomass yielding decay products	f_P
Mass N/Mass COD in biomass	i_{XB}
Mass N/Mass COD in decay products	i_{XP}

gradable organic matter (S_S) and DO (S_O), while anoxic growth also incorporates a Monod term for nitrate-nitrogen (S_{NO}). Aerobic growth of autotrophic bacteria includes Monod expressions for both ammonia (S_{NH}) and DO. Decay of heterotrophic and autotrophic bacteria are first-order expressions that continue under aerobic, anoxic, and anaerobic conditions. Ammonification of soluble organic nitrogen is modeled as a first-order process with regard to remaining substrate and biomass. Hydrolysis of slowly biodegradable organic matter (X_S) is modeled as a Monod function of the ratio of slowly biodegradable organic matter to heterotrophic bacteria, DO (S_O), and nitrate (S_{NO}). Hydrolysis of slowly degradable organic nitrogen (thereby producing ammonia) is proportional to the hydrolysis of slowly biodegradable organic matter (X_S).

Table 5-4 summarizes example values of the stoichiometric and kinetic coefficients in Activated Sludge Model No. 1 as summarized by the IAWPRC task group. These values are presented for informational purposes only and should not be viewed as typical of any wastewater or application. As discussed in Section 5.4, models must be properly calibrated prior to their use in full-scale applications. A subsequent section of this chapter discusses ex-

Table 5-3. Activated Sludge Model No. 1: Kinetic Expressions

Component → i

j	Process ↓	Process rate ρ_j, $ML^{-3}T^{-1}$
1	Aerobic growth of heterotrophs	$\hat{\mu}_H \left(\dfrac{S_S}{K_S + S_S} \right)\left(\dfrac{S_O}{K_{O,H} + S_O} \right) X_{B,H}$
2	Anoxic growth of heterotrophs	$\hat{\mu}_H \left(\dfrac{S_S}{K_S + S_S} \right)\left(\dfrac{K_{O,H}}{K_{O,H} + S_O} \right)\left(\dfrac{S_{NO}}{K_{NO} + S_{NO}} \right)\eta_g X_{B,H}$
3	Aerobic growth of autotrophs	$\hat{\mu}_A \left(\dfrac{S_{NH}}{K_{NH} + S_{NH}} \right)\left(\dfrac{S_O}{K_{O,A} + S_O} \right) X_{B,A}$
4	Decay of heterotrophs	$b_H X_{B,H}$
5	Decay of autotrophs	$b_A X_{B,A}$
6	Ammonification of soluble organic nitrogen	$k_a S_{ND} X_{B,H}$
7	Hydrolysis of entrapped organics	$k_h \dfrac{X_S / X_{B,H}}{K_X + (X_S / X_{B,H})}\left[\left(\dfrac{S_O}{K_{O,H} + S_O} \right) + \eta_h \left(\dfrac{K_{O,H}}{K_{O,H} + S_O} \right)\left(\dfrac{S_{NO}}{K_{NO} + S_{NO}} \right) \right] X_{B,H}$
8	Hydrolysis of entrapped organic nitrogen	$\rho_7 (X_{ND} / X_S)$

118

Table 5-4. Activated Sludge Model No. 1: Values of Stoichiometric and Kinetic Parameters

Symbol	Unit	Value at 20°C	Value at 10°C
Stoichiometric parameters			
Y_A	g cell COD formed/g N oxidized	0.24	0.24
Y_H	g cell COD formed/g COD oxidized	0.67	0.67
f_p	dimensionless	0.08	0.08
i_{XB}	g N/g COD in biomass	0.086	0.086
i_{XB}	g N/g COD in endogenous mass	0.06	0.06
Kinetic parameters			
$\hat{\mu}_H$	d^{-1}	6.0	3.0
K_s	g COD/m^3	20.0	20.0
$K_{O,H}$	g O$_2$/m^3	0.20	0.20
K_{NO}	g NO$_3^-$-N/m^3	0.50	0.50
b_H	d^{-1}	0.62	0.20
η_g	dimensionless	0.8	0.8
η_h		0.4	0.4
k_h	g slowly biodegradable COD/g cell COD/d	3.0	1.0
K_x	g slowly biodegradable COD/g cell COD	0.03	0.01
$\hat{\mu}_A$	d^{-1}	0.80	0.3
K_{NH}	g NH$_4^+$-N/m^3	1.0	1.0
$K_{O,A}$	g O$_2$/m^3	0.4	0.4
k_a	m^3/(g COD/d)	0.08	0.04

amples in which Activated Sludge Model No. 1 has been used and identifies actual parameter values selected.

Examination of Table 5-4 indicates values of 0.08 for f_P and 0.62 for b_H. In the unconventional approach to biomass decay described in Section 5.5.1.2 (a death-re-generation approach rather than an endogenous mass loss approach), 62 percent of the active mass dies each day (b_H = 0.62), yielding substrate for the regeneration of 38 percent, resulting in an overall net loss of 24 percent. However, now only 8 percent (f_P = 0.08) of the active mass is not biodegradable. This yields the same mass of endogenous residue as in the endogenous mass loss approach where the nonbiodegradable fraction is 0.20 and the decay coefficient is 0.24 (6). Therefore, the co-efficient values and IAWPRC equation listed in Table 2-13 yield about the same loss of active biomass and formation of X_P as obtained with Activated Sludge Model Number 1 and the coefficients for f_P and b_H listed in Table 5-4.

5.5.1.4 Model Presentation

As illustrated in Table 5-5, a matrix format is used to present the model and its structure. Listed across the top of the matrix are the components of the model; they are defined and their respective units are listed across the bottom. Microbial processes incorporated into the model are listed vertically on the left-hand side. Kinetic expressions are listed on the right-hand side of the matrix for each corresponding microbial process. Within the matrix, ·

stoichiometric coefficients are listed for each model component that is affected by a particular microbial process. Thus, the reaction rate for transformation of a particular model component is the product of the stoichiometric coefficient and the corresponding kinetic expression. For example, the reaction rate for readily biodegradable organic material (S_S) consists of aerobic growth of het-erotrophic bacteria (with stoichiometric coefficient -1/Y_H), anoxic growth of heterotrophic bacteria (with stoichiomet-ric coefficient -1/Y_H), and hydrolysis of slowly biode-gradable organic matter (with a stoichiometric coefficient of 1). Although this approach may be unusual to many, it allows the model to be presented in a particularly succinct fashion.

5.5.1.5 Examples

Experience with the application of Activated Sludge Model No. 1 to pilot or full-scale wastewater treatment systems is increasing, and the results indicate there is potential for its successful application. Several examples of the use of this model are summarized in Table 5-6. The general approach used to date involves beginning with default values for the stoichiometric and kinetic parameters in the model (e.g., those presented in Table 5-4) and then ad-justing the numerical values of these parameters to achieve an acceptable correlation of model predictions to pilot and/or full-scale data. Table 5-6 illustrates that this approach has been generally successful and indicates the types of adjustments that have been made.

Table 5.5. Activated Sludge Model No. 1 Presented in Matrix Format

Component → i

j	Process ↓	1 S_I	2 S_S	3 X_I	4 X_S	5 $X_{B,H}$	6 $X_{B,A}$	7 X_P	8 S_O	9 S_{NO}	10 S_{NH}	11 S_{ND}	12 X_{ND}	13 S_{ALK}	Process Rate, ρ_j, [ML^{-3}T^{-1}]
1	Aerobic growth of heterotrophs		$-\dfrac{1}{Y_H}$			1			$-\dfrac{1-Y_H}{Y_H}$		$-i_{XB}$			$-\dfrac{i_{XB}}{14}$	$\hat{\mu}_H\left(\dfrac{S_S}{K_S+S_S}\right)\left(\dfrac{S_O}{K_{O,H}+S_O}\right)X_{B,H}$
2	Anoxic growth of heterotrophs		$-\dfrac{1}{Y_H}$			1				$-\dfrac{1-Y_H}{2.86\,Y_H}$	$-i_{XB}$			$\dfrac{1-Y_H}{14\cdot 2.86\,Y_H}-\dfrac{i_{XB}}{14}$	$\hat{\mu}_H\left(\dfrac{S_S}{K_S+S_S}\right)\left(\dfrac{K_{O,H}}{K_{O,H}+S_O}\right)\left(\dfrac{S_{NO}}{K_{NO}+S_{NO}}\right)\eta_g X_{B,H}$
3	Aerobic growth of autotrophs						1		$-\dfrac{4.57-Y_A}{Y_A}$	$\dfrac{1}{Y_A}$	$-i_{XB}-\dfrac{1}{Y_A}$			$-\dfrac{i_{XB}}{14}-\dfrac{1}{7\,Y_A}$	$\hat{\mu}_A\left(\dfrac{S_{NH}}{K_{NH}+S_{NH}}\right)\left(\dfrac{S_O}{K_{O,A}+S_O}\right)X_{B,A}$
4	"Decay" of heterotrophs				$1-f_P$	-1		f_P					$i_{XB}-f_P i_{XP}$		$b_H X_{B,H}$
5	"Decay" of autotrophs				$1-f_P$		-1	f_P					$i_{XB}-f_P i_{XP}$		$b_A X_{B,A}$
6	Ammonification of soluble organic nitrogen										1	-1		$\dfrac{1}{14}$	$k_a S_{ND} X_{B,H}$
7	"Hydrolysis" of entrapped organics		1		-1										$k_h\dfrac{X_S/X_{B,H}}{K_x+(X_S/X_{B,H})}\left[\left(\dfrac{S_O}{K_{O,H}+S_O}\right)+\right.$ $\left.\eta_h\left(\dfrac{K_{O,H}}{K_{O,H}+S_O}\right)\left(\dfrac{S_{NO}}{K_{NO}+S_{NO}}\right)\right]X_{B,H}$
8	"Hydrolysis" of entrapped organic nitrogen											1	-1		$\rho_7(X_{ND}/X_S)$

Observed conversion rates (ML^{-3}T^{-1}) $\quad r_i=\displaystyle\sum_j v_{ij}\rho_j$

Column component definitions (M(COD)L^{-3} unless noted):
- S_I: Soluble inert organic matter, M(COD)L^{-3}
- S_S: Readily biodegradable substrate, M(COD)L^{-3}
- X_I: Particulate inert organic matter, M(COD)L^{-3}
- X_S: Slowly biodegradable substrate, M(COD)L^{-3}
- $X_{B,H}$: Active heterotrophic biomass, M(COD)L^{-3}
- $X_{B,A}$: Active autotrophic biomass, M(COD)L^{-3}
- X_P: Particulate products arising from biomass decay, M(COD)L^{-3}
- S_O: Oxygen (negative COD), M(COD)L^{-3}
- S_{NO}: Nitrate and nitrite nitrogen, M(COD)L^{-3}
- S_{NH}: $NH_4^+ + NH_3$ nitrogen, M(COD)L^{-3}
- S_{ND}: Soluble biodegradable organic nitrogen, M(COD)L^{-3}
- X_{ND}: Particulate biodegradable organic nitrogen, M(COD)L^{-3}
- S_{ALK}: Alkalinity-molar units

Stoichiometric Parameters:
Heterotrophic yield: Y_H
Autotrophic yield: Y_A
Fraction of biomass yielding particulate products: f_P
Mass N/Mass COD in biomass: i_{XB}
Mass N/Mass COD in products from biomass: i_{xp}

Kinetic Parameters:
Heterotrophic growth and decay: $\hat{\mu}_H$, K_s, $K_{O,H}$, K_{NO}, b_H
Autotrophic growth and decay: $\hat{\mu}_A$, K_{NH}, $K_{O,A}$, b_A
Correction factor for anoxic growth of heterotrophs: η_g
Ammonification: k_a
Hydrolysis: k_h, K_x
Correction factor for anoxic hydrolysis: η_h

120

Table 5-6. Examples: Use of IAWPRC Activated Sludge Model No. 1

Facility	Application	Adjustments Required	Comments	Reference
Six Municipal Facilities	Prediction of process oxygen requirements		Only minor modifications required for each facility	7
DSM Chemicals, Augusta, GA	Design pilot plant study, design full-scale facility	Y_H, b_H, η_g, η_h, $\hat{\mu}_A$, b_A	Extensive wastewater characterization, and analysis of pilot plant data. Industrial wastewater.	8
Valenton, France; CMRL Pilot Plant, France; Henin, France;	Process evaluations and designs	η_g, $\hat{\mu}_A$	Model modified to incorporate adsorption of readily degradable colloids.	9
Flauil, Switzerland; Zurich, Switzerland; Dietikar, Switzerland;	Measurement of process parameters, process analysis	$\hat{\mu}_H$, K_S, b_H	Used first-order rate expression for hydrolysis of Xs. Did not evaluate nitrification.	10
Zurich, Switzerland	Process Analysis	$\hat{\mu}_H$, Y_H, K_S, $\hat{\mu}_A$	Used first-order rate expression for hydrolysis of Xs.	11
Renton, WA; Santa Fe, NM	Process oxygen requirement distribution	None	None	12

The results presented in Table 5-6 suggest that site-specific adjustments may be required most often for the maximum specific growth rates of the heterotrophic ($\hat{\mu}_H$) and autotrophic ($\hat{\mu}_A$) bacteria. Adjustments in $\hat{\mu}_H$ have been required to produce acceptable oxygen uptake rate/DO profiles along plug flow biological reactors. Adjustments in $\hat{\mu}_A$ are necessary because of the effect of temperature and the presence of inhibitory materials on the nitrifying bacteria. The individual references should be consulted concerning the magnitude of the required adjustment; however, adjustments of plus or minus 50 percent are not uncommon. Also, note the effect of parameter interaction. For example, an adjustment in $\hat{\mu}_H$ may also necessitate a corresponding adjustment in K_s. While ongoing application of the model and detailed review of the results is necessary to develop additional insight, the results to date are certainly encouraging.

Wastewater characteristics must also be assessed appropriately to allow the model to predict system responses accurately. The case histories referenced in Table 5-6 also present insight into this issue.

5.5.2 Fixed Growth Models

In many respects, the modeling of fixed growth systems is more difficult than the modeling of suspended growth systems. The increased difficulty results from at least two factors. One is the effect of diffusional resistance on the substrate removal rate in such systems, as discussed in Chapters 3 and 4. To accurately predict the performance of fixed growth systems, the effects of diffusional resistance must be incorporated into the model. These additional factors do not need to be considered explicitly in suspended growth models.

A second factor is the heterogeneous and relatively poorly characterized conditions that occur within many fixed growth reactors. For example, in rotating biological contactors (RBCs) the growth media (and, consequently, also the biological film) is alternately submerged in the completely mixed reactor basin and exposed to the atmosphere. Perhaps the most heterogeneous environment is that which exists in a trickling filter. In most trickling filters, influent flow is dosed on a periodic basis using a rotating distributor. Consequently, the media is alternately wetted and then exposed directly to the atmosphere. Similarly, the hydrodynamic conditions within the trickling filter are poorly characterized (13). Since liquid flow across the trickling filter media is not sheet flow (i.e., not all the trickling filter media surface is fully wetted), characterization for modeling is complicated.

Modeling is also hampered by fundamental gaps in knowledge, such as the lack of information about the factors that affect the fate of particulate matter in biofilms. While it is known that certain biofilms will entrap and subsequently metabolize some of the particulate matter in an applied wastewater (14), and that entrapped particulates will reduce the activity of the biomass as they displace active biomass from the biofilm, the mechanisms of attachment and metabolism for particulate matter re-

main poorly understood. Additional research is needed to resolve these and other issues so that more mechanistically correct models can be developed.

In spite of these difficulties, research continues on the development of models of fixed film processes that accurately predict process performance. An IAWPRC task force, similar to that which developed Activated Sludge Model No. 1, is currently working on the development of a general purpose biofilm model. In addition, researchers are working on the development of models to predict the nitrification performance of fixed film processes, either singly or in combination with carbon oxidation. Two examples include the RBC model developed by Gujer and Boller (15) and the trickling filter model developed by Logan, as modified and extended for nitrification by Parker et al. (16).

5.5.3 Available Computer Programs for Modeling

Several microcomputer-based programs have been developed to facilitate use of the process models described above. Table 5-7 summarizes several features of these programs for suspended growth systems and provides contacts for the software. The information is intended to

Table 5-7. Example Nitrification and Denitrification Mathematical Modeling Microcomputer Packages

Name	Features	Contact
SSSP	Direct implementation of Activated Sludge Model No. 1. Public domain	C.P.L. Grady, Jr. Environmental Syst. Engr. Rich Environmental Research Laboratory Clemson Research Park Clemson, SC 29634
ASIM	Flexible modeling tool that can incorporate Activated Sludge Model No. 1 or more complicated model	Dr. Willi Gujer Abt. Ingenieurwissen Schaften Überlandstrasse 13 EAWAG CH-8600 Dubendorf Switzerland
EFOR	Activated Sludge Model No. 1 plus clarifier model	Mr. Jan Peterson EFOR I. Kruger AS Gladsaxevej 363 DK-2860 Soborg Denmark
GPS-X	General purpose simulator, Activated Sludge Model No. 1 plus clarifier	Hydromantis, Inc. 1685 Main St. West Suite 302 Hamilton Ontario L8S 1G5

illustrate the availability of such tools, not to serve as an endorsement of any particular commercial product.

5.6 Example: Analysis of a Single-Sludge Nitrification/Denitrification System

This section demonstrates the use of the IAWPRC Activated Sludge Model No. 1 to analyze and optimize a single-sludge nitrification/denitrification system, using the influent wastewater characteristics presented in Chapter 2 along with information on the single-sludge nitrification/denitrification systems developed in Chapter 8. Thus, the results of this analysis can be compared to data presented in Chapter 8.

5.6.1 Development of Wastewater Characteristics

The initial step in the application of the IAWPRC Activated Sludge Model No. 1 is development of the specific influent wastewater characteristics required by the model (see Table 5-1). (See Chapter 2 for a more detailed description of the selected wastewater treatment plants and the biological process influent wastewater characteristics.)

First, consider the simple Plant A as described in Figure 2-5 and Table 2-15, and calculate the particulate inert organic matter (X_I). According to Table 2-11, the particulate COD is 180 mg/L and it is 30 percent nonbiodegradable; that is 0.3 x 180 mg/L, or 54 mg/L. For simplicity, use X_I equal to 55 mg/L (expressed as COD). Assuming that the slowly biodegradable substrate (X_S) is equal to the biodegradable particulate COD (see Section 5.5.1.1), X_S may be calculated as the difference between the particulate COD and the particulate inert organic matter, which is 180 – 55 or 125 mg/L (expressed as COD). According to Tables 2-11 and 2-15, the soluble COD is 100 mg/L and the soluble nonbiodegradable organic matter is roughly 20 mg/L, expressed as COD; consequently, it is assumed that the readily biodegradable substrate (S_S) is calculated as 100 – 20 or 80 mg/L (expressed as COD).

According to Table 2-11, the total influent soluble nitrogen is 24 mg-N/L. Assume that the ammonia-nitrogen (S_{NH}) concentration is 20 mg-N/L and the influent nitrate plus nitrite (S_{NO}) concentration is zero. Neglecting nonbiodegradable soluble organic nitrogen, the soluble degradable organic nitrogen (S_{ND}) would be the difference between the influent soluble nitrogen and the ammonia-nitrogen concentration of 24 – 20, or 4 mg-N/L. The total influent particulate nitrogen concentration is 6 mg/L. Assuming that 1 mg/L of this concentration is nonbiodegradable, the particulate biodegradable organic nitrogen (X_{ND}) is 5 mg-N/L. The influent alkalinity is 120 mg/L as $CaCO_3$. Since there are 50 mg/L of alkalinity as $CaCO_3$ per mM, this corresponds to 2.4 mM of bicarbonate alkalinity. In summary, the biological reactor influent wastewater charac-

teristics for the simple wastewater are:

X_I = 55 mg/L as COD
S_S = 80 mg/L as COD
X_S = 125 mg/L as COD
S_{NO} = 0 mg-N/L
S_{NH} = 20 mg-N/L
S_{ND} = 4 mg-N/L
X_{ND} = 5 mg-N/L
S_{ALK} = 2.4 mM

A similar analysis of the mass balance information from the more complex wastewater treatment Plant B (Figure 2-6 and Table 2-16) yields the following estimates for the biological reactor influent (primary effluent) wastewater characteristics:

X_I = 29 mg/L as COD
S_S = 86 mg/L as COD
X_S = 52 mg/L as COD
S_{NO} = 0 mg-N/L
S_{NH} = 22.4 mg-N/L
S_{ND} = 4 mg-N/L
X_{ND} = 2 mg-N/L
S_{ALK} = 2.4 mM

For both cases, the influent active heterotrophic biomass ($X_{B,H}$) and active autotrophic biomass ($X_{B,A}$) are assumed to be zero.

5.6.2 Detailed Analysis

As a further illustration of the use of the model, a design will be developed for a 18,925 m³/d (5 mgd) average flow facility with a peak month flow of 28,388 m³/d (7.5 mgd). The more complex wastewater treatment Plant B is considered, and the facility is designed to meet a 10 mg-N/L total nitrogen standard. The design provides an anoxic volume of 2,914 m³ (0.77 MG) and an aerobic volume of 5,450 m³ (1.44 MG). The anoxic and aerobic zones are each divided into three cells in series. The return activated sludge rate is 72 percent of the peak month flow, and the internal recycle is 160 percent of the peak month flow. Oxygen transfer is provided by six slow-speed surface mechanical aerators, each operating at 50 hp. A uniform aeration pattern is proposed. Results from this analysis can be compared to an alternative design approach shown for Example 1 in Chapter 8 (Section 8.5.3).

The public domain computer program SSSP was used to carry out the analysis in this example. Table 5-8 illustrates the computer input and output. Critical inputs include a maximum specific growth rate for autotrophs of 0.45 d⁻¹ (as was used for the example in Chapter 8), a total mean cell residence time (MCRT), or solids retention time of 12.1 d (giving an aerobic MCRT of 9.4 d), and a K_La in the aerobic section of 125 d⁻¹ (based on uniform aeration

with the six 50 hp aerators). The design temperature would be 15°C (59°F). The results indicate the following:

- The calculated MLVSS (mixed liquor volatile suspended solids) in the aeration basin effluent is about 3,000 mg/L as COD, which corresponds to a concentration of approximately 2,110 mg/L when expressed as VSS (calculated using 1.42 mg COD/mg VSS). This exceeds the design MLVSS concentration of 1,890 mg/L (expressed as VSS), as presented in Chapter 8, and indicates a slightly higher solids loading on the secondary clarifier.

- The predicted effluent total inorganic nitrogen concentration (NH_4^+-N plus NO_x-N) is 7.7 mg/L, consisting of 0.5 mg/L for NH_4^+-N and 7.2 mg/L for NO_x-N. Thus, the effluent organic nitrogen concentration could be as much as 2.3 mg/L (10 mg/L minus 7.7 mg/L) and the effluent total nitrogen concentration would be less than 10 mg/L. Again, the viability of achieving this level of performance must be carefully evaluated.

- The NO_x-N concentration profile through the anoxic zone (reactors 1, 2, and 3) is 3.0, 1.8, and 1.2 mg/L. This indicates that more than enough nitrate is returned to this zone to ensure that the denitrification potential is fully utilized.

- The ammonia-nitrogen profile through the aerobic zone (reactors 4, 5, and 6) indicates that the aerobic zone is fully utilized. The fact that the ammonia-nitrogen concentrations is just reduced to 0.5 mg/L in the last stage of the aerobic zone (reactor 6) indicates that the entire aerobic zone must be used for nitrification.

- Denitrification is predicted to occur in both the anoxic and aerobic zones. Based on the unit rates of denitrification, it can be calculated that the total amount of nitrogen denitrified is 16.8 mg/L. Of this total, 19 percent is predicted to occur in the aerobic zone (reactors 4, 5, and 6). Denitrification can occur in the aerobic zone (at least according to this model) as a result of the use of Monod kinetics for denitrification. Denitrification is predicted to occur in an aerated zone when the DO concentration is relatively low and the NO_x-N concentration is relatively high.

- The DO concentration profile in the aerobic zone (reactors 4, 5, and 6) indicates that the placement of aeration capacity is reasonable. A DO concentration of 1.4 mg/L is predicted for the first stage of the aerobic zone (reactor 4), and the DO concentration is predicted to be 3.6 mg/L in the effluent from the aerobic zone (reactor 6). A reduced DO concentration in reactor 6 would further optimize process performance.

- Supplemental alkalinity is required, as indicated by a reactor effluent alkalinity of 0.3 mM. This corresponds to an effluent alkalinity of 15 mg/L as $CaCO_3$. Supple-

123

Table 5-8. SSSP Input Parameters and Output for Example Problem

Parameter	Kinetic and Stoichiometric Parameters	
	Heterotrophic Organisms	Autotrophic Organisms
μ_{max}, d^{-1}	4.0	0.45
k_s COD, g COD/m^3	10.0	
k_s NH$_4$-N, g N/m^3		1.0
k_s O$_2$, g O$_2$/m^3	0.55	0.55
Yield, g/g	0.67	0.24
b decay, d^{-1}	0.62	0.05
Anoxic growth factor	0.80	
k_s NO$_3$, g N/m^3	0.20	
Hydrolysis rate, d^{-1}	2.20	
Hydrolysis saturation ratio, g COD/g COD	0.15	
Anoxic hydrolysis factor	0.40	
Ammonification, m^3/g COD/d	0.16	
Fraction of particulate products, g COD/g COD	0.08	
N in biomass, g N/g COD	0.086	
N in part. prod., g N/g COD	0.06	
O$_2$ saturation concentration, g O$_2$/m^3	9.0	

Reactor Specification	Process Configuration and Flow Distribution[a]					
	Reactor Number					
	1	2	3	4	5	6
Reactor volume, m^3	971	971	971	1817	1817	1817
Feed fraction, 0 to 1	1.0	0	0	0	0	0
Mass trans. coeff. for O$_2$, d^{-1}	0	0	0	125.0	125.0	125.0
Recycle input, m^3/d	20439	0	0	0	0	0
Recirculation input, m^3/d	45421	0	0	0	0	0
Recirculation originated from reactor	6					

Constituents	Steady-State Solution[b]						
	Feed	Location					
		1	2	3	4	5	6
Heterotrophic organisms, g COD/m^3	0.0	951.7	952.6	950.1	949.6	949.7	950.0
Autotrophic organisms, g COD/m^3	0.0	140.7	140.7	140.6	141.2	141.6	141.9
Particulate products, g COD/m^3	0.0	582.6	583.1	583.5	584.5	585.4	586.3
Inert particulates, g COD/m^3	29.0	1191.0	1191.0	1191.0	1191.0	1191.0	1191.0
Particulate organics, g COD/m^3	52.0	142.8	144.5	146.5	140.5	134.0	127.1
Soluble organics, g COD/m^3	86.0	9.7	3.3	1.6	1.9	2.0	2.0
Soluble ammonia N, g N/m^3	22.4	6.8	6.8	6.9	4.1	1.7	0.5
Soluble nitrate/nitrite N, g N/m^3	0.0	3.0	1.8	1.2	3.7	5.9	7.2
Soluble organic N, g N/m^3	4.0	0.7	0.4	0.3	0.4	0.4	0.5
Biodegrad. part. organic N, g N/m^3	2.0	10.5	10.7	11.0	10.6	10.3	9.8
Oxygen, g O$_2$/m^3	0.0	0.1	0.0	0.0	1.4	2.1	3.6
Alkalinity, mole/m^3	2.4	1.1	1.2	1.2	0.8	0.5	0.3
MLVSS, g COD/m^3		3008.8	3011.8	3011.7	3006.7	3001.7	2996.2
O$_2$ consumed, g O$_2$/m^3/d		163.1	4.8	0.2	875.6	819.6	601.0
Nitrate consumed, g NO$_3^-$-N/m^3/d		221.2	115.4	60.5	22.1	16.9	11.1

[a] Number of Reactors = 6, Solids Retention Time = 12.1, Average Flow Rate = 28,388 m^3/d.
[b] Warnings: The average was taken of the k_s O$_2$ parameters. The alkalinity in reactors 4, 5, and 6 is below the 1 mole/m^3 required to sustain uninhibited biological growth.

mental alkalinity of about 35 mg/L as $CaCO_3$ would be required.

The model could be used in a number of ways to optimize the process. First, the variation in process oxygen requirements through the aerobic zone (reactors 4, 5, and 6) suggests the use of tapered rather than uniform aeration. Several variations were investigated using the model. A 20-percent increase in the horsepower in the first stage of the aerobic zone (from 50 to 60 hp) was simulated by increasing the K_La from 125 d^{-1} to 150 d^{-1}. This resulted in a DO concentration in this cell of 2.0 mg/L, which led to less predicted denitrification in the aerobic zone and an increased process effluent NO_x-N concentration. A reduction in the aerator horsepower in reactor 6 was simulated by reducing K_La to 60 d^{-1}. This resulted in a reduction in the DO concentration to 2.9 mg/L (from the previous value of 3.6 mg/L) and a reduction in the effluent NO_x-N to 6.3 mg/L due to increased denitrification in the aerobic zone. However, based on the data available, the use of two-speed aerators with different sizes might be appropriate. Sixty hp in the first aerobic zone (reactor 4), 50 hp in the second aerobic zone (reactor 5), and 40 hp in the third aerobic zone (reactor 6) would represent a reasonable distribution.

If diurnal flow and load data are available, a more refined analysis of the aeration system could be conducted. Such an analysis would facilitate more precise sizing of the aeration system to meet peak oxygen transfer requirements while not providing excess oxygen transfer capacity. Such an analysis would also allow precise sizing of the biological reactor to treat peak ammonia-nitrogen loadings and avoid ammonia breakthrough. This, in fact, is one of the primary benefits of the use of a model such as Activated Sludge Model No. 1. Because it is a dynamic model and is capable of quantifying such effects, it allows more precise sizing of the system to adequately treat the range of loads applied to it.

The capability to predict denitrification within the aerobic zone is an interesting aspect of this model. However, such predictions must be compared with full-scale results as part of the model calibration and verification exercise described in Section 5.3.

5.6.3 Alternative Comparison

The final use of Activated Sludge Model No. 1, which will be illustrated, involves a comparison of reactor sizes required for the two plant types identified in Chapter 2 and for designs intended to meet effluent total nitrogen limits of 10 and 5 mg/L. The comparison is based on the wastewater characteristics presented above. Results can also be compared to design examples 1 and 2 in Section 8.5 and the design detailed in Table 5-8 to achieve a TN effluent limit of 10 mg/L.

Comparable designs were developed by maintaining MLVSS concentrations (expressed as VSS) of about 2,300 mg/L. An aerobic solids retention time of 9.4 d was maintained for the 10 mg-N/L effluent total nitrogen cases and a more conservative aerobic solids retention time of 11.7 d was used for the 5 mg-N/L effluent total nitrogen cases. To ensure consistency in the comparison, a uniform DO concentration of 2 mg/L was maintained through the aerobic reactors.

The results of the analysis are presented in Table 5-9. Process designs consisting of a single anoxic and a single aerobic zone are utilized for the 10 mg/L effluent total nitrogen cases, while dual anoxic zone systems are util-

Table 5-9. Comparison of Process Designs for Complex and Simple Wastewater Treatment Plants and Effluent Total Nitrogen of 10 and 5 mg-N/L

Parameter	TN = 10 Complex	TN = 10 Simple	TN = 5 Complex	TN = 5 Simple
HRT, hr[a]				
First anoxic	2.5	2.5	2.4	2.4
First aerobic	4.6	8.1	5.3	9.9
Second anoxic	—	—	2.7	2.7
Second aerobic	—	—	0.5	0.5
TOTAL	7.1	10.6	10.9	15.5
MCRT, d				
Total	14.4	12.1	24.0	18.3
First aerobic	9.4	9.4	11.7	11.7
MLVSS, mg/L	2,312	2,360	2,384	2,327
Recirculation input, %[b]	160	160	400	400
RAS, %[b]	72	72	72	72
NH_4^+-N, mg-N/L	0.1	0.2	0.1	0.2
NO_x-N, mg-N/L	7.1	4.9	3.6	0.9
Denitrification, mg-N/L	17.5	18.3	21.3	22.4
Aerobic zone denitrification, %	21	38	17	31
First anoxic zone effluent NO_x-N, mg-N/L	1.8	0.0	2.1	0.2
Effluent alkalinity, mg/L as $CaCO_3$	15	30	30	40

[a] Based on maximum month flow, without recycle.
[b] Percentage of maximum month flow.

ized for the 5 mg/L effluent total nitrogen cases. The required hydraulic residence times (HRTs) were 7.1-15.5 hr. The HRT increased by about 4-5 hr as the effluent total nitrogen limit was reduced from 10 to 5 mg/L. The difference in total HRT between Plants A and B was similar.

The results also indicate differences in performance characteristics. For example, effluent NO_x-N concentrations are lower for the simple wastewater treatment plant examples than for the complex treatment plant examples. The model predicts a greater increase in denitrification for the simpler Plant A than for the complex Plant B. Interestingly, increased denitrification is predicted to occur in the aerobic portion of the reactor (17 to 21 percent for the complex wastewater treatment Plant B as compared to 31 to 38 percent for the simpler wastewater treatment Plant A). In addition, essentially complete denitrification is predicted to occur in the first-anoxic-zone for the simpler Plant A (as indicated by low first-anoxic-zone effluent NO_x-N concentrations), while denitrification is not complete for the complex Plant B. Differences in effluent alkalinity are predicted for the four cases considered.

The proposed design that would allow the complex Plant B to meet an effluent total nitrogen of 5 mg/L predicts an effluent total inorganic nitrogen concentration (NO_x-N plus NH_4^+-N) of 3.7 mg/L. Alternatives to lower this value, and thus provide a greater margin of safety relative to the 5 mg/L effluent total nitrogen limit, were investigated. Neither an increase in the size of the first anoxic zone nor an increase in the internal recycle rate resulted in significant changes in the effluent total inorganic nitrogen concentration. However, the effluent total inorganic nitrogen was significantly reduced by converting the last portion of the first aerobic zone to anoxic operation. This effectively reduced the size of the first aerobic zone, but significantly increased the size of the second anoxic zone. These modifications could result in a reduction in the effluent total inorganic nitrogen concentration to about 2 mg/L.

5.7 References

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Chapter 6
Design Considerations for Biological Nitrification Processes

6.1 Introduction

Biological nitrification in municipal wastewater treatment is particularly applicable to those cases where an ammonia removal requirement exists, without the need for complete nitrogen removal. Biological nitrification is also a key step in the biological nitrification-denitrification approach to nitrogen removal.

This chapter deals with design aspects of specific nitrification process systems and the influence of various factors on nitrification kinetics in these systems. Methods for the design of nitrification reactors are presented, together with design examples and operating and performance information from pilot and full-scale plants.

6.2 Classification of Nitrification Processes

In Section 3.3.6 (Effect of Feed Organic Carbon to Nitrogen Ratio), the feed C:N was shown to be a critical factor affecting the design of nitrification systems. Previous design documents categorized nitrification systems according to the degree of separation of the CBOD removal and nitrification processes, often using the C:N as a benchmark [1,2]. A nitrification system was categorized as a combined or single-sludge carbonaceous oxidation-nitrification system if the $CBOD_5$:TKN was greater than a certain arbitrary value, typically 4 or 5. Below the selected value, the system was considered a separate, or two-sludge, system. In this manual, nitrification systems are not strictly categorized in this fashion, although systems at higher and lower feed $CBOD_5$ concentrations are compared in Section 6.3, and the importance of the feed C:N is illustrated through completion of design examples at higher (Section 6.4.2.2) and lower (Section 6.4.3.2) ratio values.

Biological nitrification reactors can be classified according to the nature of their biological growth. Activated sludge systems suspend biological solids in a mixed liquor by some mixing mechanism; these are termed suspended growth reactors. Units in which growth occurs on or within a solid medium are termed attached growth, supported growth or fixed film reactors. Certain bioreactors contain microbial films in suspension, resulting from the addition of very fine inert or active particles to provide microbial growth sites. An example is the powdered activated carbon (PAC) activated sludge reactor. Such reactors are considered suspended growth reactors since the kinetic reactions are well described by equations appropriate for suspended growth [3]. In certain nitrification systems, the growth of both suspended and attached biomass is promoted in the reactor. An example of this combined growth configuration involves the suspension of highly porous plastic foam particles in the reactor; attached growth occurs in and around the plastic media while suspended growth occurs in the liquid phase. Information pertaining to these systems is presented in Section 6.6.

There are many different configurations of suspended, attached, and combined growth reactors for nitrification. New configurations are frequently introduced. A list of alternatives is presented on Figure 6-1; this does not include reactors designed to achieve both nitrification and partial denitrification such as Passveer ditch type systems, or systems designed to achieve both nitrification and biological phosphorus removal. Systems designed to achieve total nitrogen removal (i.e., nitrification and denitrification) are discussed in Chapter 8. Descriptive and design information pertaining to suspended, attached and combined growth nitrification reactors is provided in Sections 6.4, 6.5 and 6.6, respectively. Design considerations to incorporate phosphorus removal into suspended growth nitrification systems are presented in Section 6.4.10.2.

6.3 Comparison of Nitrification Systems at Higher and Lower Carbonaceous Feed Concentration

Representative nitrification systems are classified in Table 6-1 according to the nature of the biological growth in the process reactor(s). Included are systems with both higher and lower feed $CBOD_5$:TKN values. Table 6-1 also shows the distribution of total oxygen demand in the process between carbonaceous sources ($CBOD_5$) and nitrogenous sources. It can be seen that in systems with lower $CBOD_5$:TKN, the proportion of nitrogenous oxygen demand (NOD) is at least 70 percent of the total. In systems
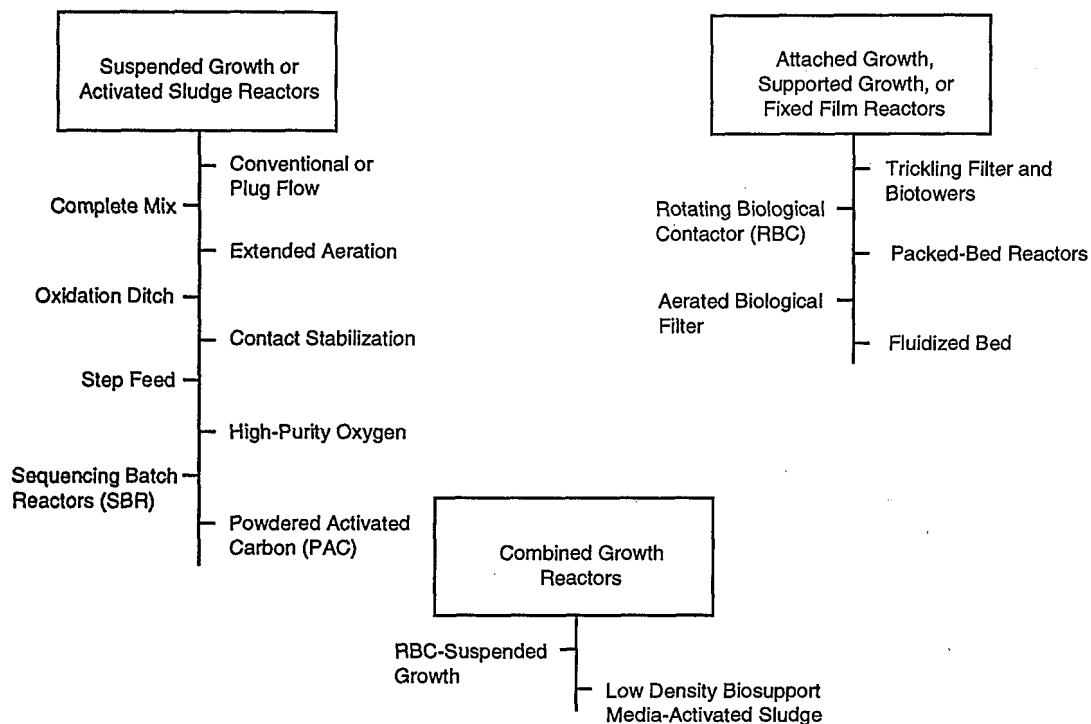
Figure 6-1. A listing of the majority of reactor configurations available for nitrification.

Table 6-1. Classification of Nitrification Facilities (Adapted in Part from Reference 1)

Location	Nature of Biological Growth	Reactor Feed CBOD$_5$:TKN	Oxygen Demand Distribution (Percent)		Reference
			CBOD$_5$	NOD	
Manassas, VA	Suspended	1.2	20	80	4
South Bend, IN	Suspended	1.8	28	72	5
Fitchburg, MA	Suspended	1.0	18	82	6
Whittier Narrows, LACSD, CA	Suspended	6.6	61	39	7
Cincinnati, OH	Suspended	7.2	61	39	8
Allentown, PA	Attached	1.9	30	70	9
Stockton, CA	Attached	5.3	54	46	10

130

with higher ratios, the proportion of nitrogenous oxygen demand is lower than 50 percent.

As was shown in Section 3.3.6 (Effect of Feed Organic Carbon to Nitrogen Ratio), the design solids retention time, θ_c^d, is inversely proportional to the rate of substrate removal (Equation 3-18). The rate of substrate removal, q_H, is directly proportional to the feed total CBOD, S_o, according to Equation 3-19:

$$q_H = \frac{S_o - S_1}{X_1 t} \qquad (3\text{-}19)$$

Reducing the feed total CBOD only will reduce q_H, effectively increasing the design solids retention time. One way to reduce the feed CBOD is to place an organic carbon removal step ahead of the nitrification step. This separate two-stage configuration yields a nitrification process with a lower carbonaceous feed concentration, reducing the food available for growth of heterotrophic bacteria in the second stage. Practical reactor nitrifying biomass concentrations can be maintained at long solids retention times (15 to 25 days), with reasonable hydraulic retention times.

A procedure for reducing the substrate removal rate, but without separating the carbon oxidation and nitrification processes, is to increase the biological solids in the system. This single-stage configuration uses higher concentrations of biological solids (i.e., the MLVSS in a suspended growth reactor) or increases the volume of the reactor, while maintaining the concentration of biological solids at the same level.

The initial development of a suspended growth nitrification system with a low feed $CBOD_5$ concentration was oriented to the isolation of the carbonaceous removal and nitrification steps so that each could be separately controlled and optimized (11). By placing a carbon removal step ahead of the nitrification step, the second-stage sludge would contain a higher percentage of nitrifying organisms per unit of MLVSS than found in a system receiving a high feed $CBOD_5$. A higher percentage of nitrifying organisms makes it easier to increase the reactor solids retention time thus providing a less temperature sensitive process configuration. The first applications of two-stage suspended growth nitrification systems were in the northern portion of the United States where low liquid temperatures (i.e., less than 10°C [50°C]) were experienced in the wintertime. Such systems were later applied in moderate climates such as Florida and California (4).

There have been difficulties maintaining a sufficient nitrifying sludge inventory in low $CBOD_5$ suspended growth nitrification systems (12–14). Typically, effluent solids were observed to fluctuate between 10 and 50 mg/L, composed primarily of dispersed solids that were not captured in the secondary clarifier. It has been suggested that a high fraction of the mixed liquor must be heterotrophic in order to maintain good bioflocculation in nitrification systems (5). An unstable condition will exist in these low feed $CBOD_5$ systems if the effluent solids exceed the net yield of organisms grown in the system and captured by the clarifier. Several remedies are available. At some locations, solids from the upstream carbonaceous oxidation suspended growth reactors were periodically transferred to the nitrification stage to maintain an adequate solids inventory. In other cases, first-stage pretreatment configurations were chosen such that lower $CBOD_5$ removals were accomplished when compared to suspended growth systems. This raised the $CBOD_5$ feed to the nitrification stage, increasing the synthesis of heterotrophic bacteria and improving the settleability of the mixed liquor solids. Examples include primary clarification with chemical addition (15) and the use of attached growth reactors (e.g., trickling filter) for partial carbonaceous removal. Further information concerning the coupling of attached and suspended growth reactors for sequential carbonaceous removal and nitrification is presented in Section 6.6. In another case, 10 percent of the primary effluent was bypassed around the carbonaceous removal step to the nitrification step to overcome the low solids synthesis issue (16). The amount of primary effluent bypassed can be varied in order to control the solids retention time according to a procedure presented elsewhere (17). Recirculation of filter backwash solids to the nitrifying system also is effective and is the method of choice at a number of facilities with effluent filters.

6.3.1 Pretreatment for Removal of Carbonaceous Material and Inhibitory Compounds

Pretreatment alternatives provide varying degrees of organic carbon removal ahead of the nitrification step. As discussed earlier, a high degree of organic carbon removal through pretreatment will lead to the highest ammonium oxidation rates per unit of MLVSS in the downstream nitrification stage. This implies that reactor size will diminish with increasing degrees of carbon removal in the pretreatment stage. But low feed carbon levels can cause difficulties in separate stage suspended growth nitrification systems, as discussed in Section 6.3. Conversely, low levels of organics in the influent to attached growth reactors can be advantageous because of reduced competition on the media surface between heterotrophic and nitrifying organisms and because the low synthesis of solids results in very low levels of solids in the effluent. In some cases this can eliminate the need for a clarification step, especially if followed by effluent filtration or some other downstream treatment process such as denitrification.

The selection of a pretreatment step to reduce the feed $CBOD_5$ to the nitrification stage can also be beneficial in protecting the nitrifiers against inhibitory compounds present in the wastewater as a result of industrial discharges. The types of inhibitory compounds removed will vary, depending on the specific unit operations employed in the pretreatment step. Biological pretreatment alternatives

131

provide a degree of protection against both organic and heavy metal inhibitors. An exception would be organics that are difficult to degrade or are resistant to biological oxidation, such as, respectively, the solvents perchloroethylene and trichloroethylene which have been identified as inhibitors to nitrification (5,18). Lime or metal salt chemical treatment is one of the most effective processes for removal of a wide range of metals (19).

Federal and state pretreatment regulations are directed toward eliminating pollutants that are incompatible with the operation and performance of municipal treatment plants. An effective industrial pretreatment program should preclude concern for issues regarding inhibition of the biological processes at a treatment plant. If there is a problem with inhibitory levels of organics or metals, the source should be located. If it cannot be eliminated through some pollution prevention, waste minimization techniques, then pretreatment will be necessary before discharge.

Procedures have been developed to screen for nitrifying inhibitors in the wastewater and to assess the effectiveness of various pretreatment alternatives at reducing inhibition (20,21). These might be considered when there are major industrial users in the system, and the potential exists for the discharge of problem compounds. Perhaps the simplest screening procedure involves batch oxygen uptake tests using a respirometer to measure oxygen utilization (21). Composite wastewater samples are subjected to various pretreatments (e.g., alum or powdered activated carbon via a jar test procedure or to biological oxidation by batch aeration). Each treated sample is then split and placed into two respirometers. One respirometer is used as a non-nitrifying control by treatment with a nitrification inhibitor such as Allylthiourea or N-Serve (2-chloro-6 trichloromethyl pyridine). Both respirometers are inoculated with a small amount of mixed liquor from a nitrifying activated sludge plant. Alternatively, a seed acclimated to the wastewater in question may provide more meaningful results where the inhibiting compound is biodegradable; the compound may be removed either in a batch pretreatment evaluation or by the heterotrophic component of an acclimated nitrifying sludge used in the respirometer study. Differences between the oxygen used in the control and in the seeded sample can be used to establish batch nitrification rates. At the end of the test, the respirometer contents are sampled and analyzed for the nitrogen species to confirm whether nitrification took place in the inoculated samples as well as to check the control. The adequacy of the seed used can also be checked by running an inoculated, but uninhibited, sample known to contain ammonia and organics, but no inhibitors.

The batch nitrification rates can be examined to determine the most suitable pretreatment technique among the options examined. Often, some of the pretreatment techniques will result in little or no nitrification in the inoculated

sample, indicating inadequate removal of the inhibitor(s). In other cases, the pretreatment techniques will allow vigorous nitrification in the sample indicating good removal of the inhibitor(s). The particular pretreatment technique that is effective may also indicate the type of inhibitor that is interfering with nitrification and may permit identification and elimination of the source to the system. For instance, if lime treatment is effective, the problem may be a heavy metal that can be precipitated by lime. Alternatively, if biological oxidation is ineffective but activated carbon treatment allows nitrification to proceed, then a nonbiodegradable organic is suspect. Subsequent specific analyses can then be run in the identified category of compounds. If the inhibitors cannot be eliminated by a source control program, often a pilot study of the process identified by the bench scale procedure can be justified to confirm the process selection. Pilot studies also have value in determining the ability of the nitrifiers to adapt to the toxicants, something the batch test with an unacclimated seed is not capable of doing. Note that the use of chemical addition as a pretreatment step to the nitrification stage may cause significant changes in alkalinity and pH. Normally this pretreatment approach is selected in order to achieve a degree of phosphorus removal as well. The effects of chemical addition on alkalinity and pH are discussed in Chapter 2 (Section 2.7.2).

6.4 Nitrification in Suspended Growth Reactors

Suspended growth systems were commonly used in England to obtain dependable nitrification long before their use became widespread in the United States. Early U.S. conventional activated sludge plants often nitrified in the warmer months of the year or if they were underloaded. Nitrification became unpopular because of the additional aeration power cost and the propensity of some sludges to float in the sedimentation tank as a result of denitrification, and it was questioned whether the process was worth the added expense in many cases (22). As a consequence, ways and means were sought to prevent nitrification, rather than to encourage it, by increasing organic loading, using tapered aeration, or picking modifications of the process which were less favorable for nitrification. This early experience with the process may have led to some early uncertainty about its reliability.

At least eight suspended growth reactor configurations can be designed to incorporate nitrification (Figure 6-1), all of which are modifications of the activated sludge process. It may be appropriate to classify plug flow, complete mix, contact stabilization, oxidation ditch, and sequencing batch reactors as aeration basin configurations and the remaining configurations as process modifications to these reactors. Simplified schematic representations of four of the systems are presented in Figure 6-2. Descriptive information and design and performance information
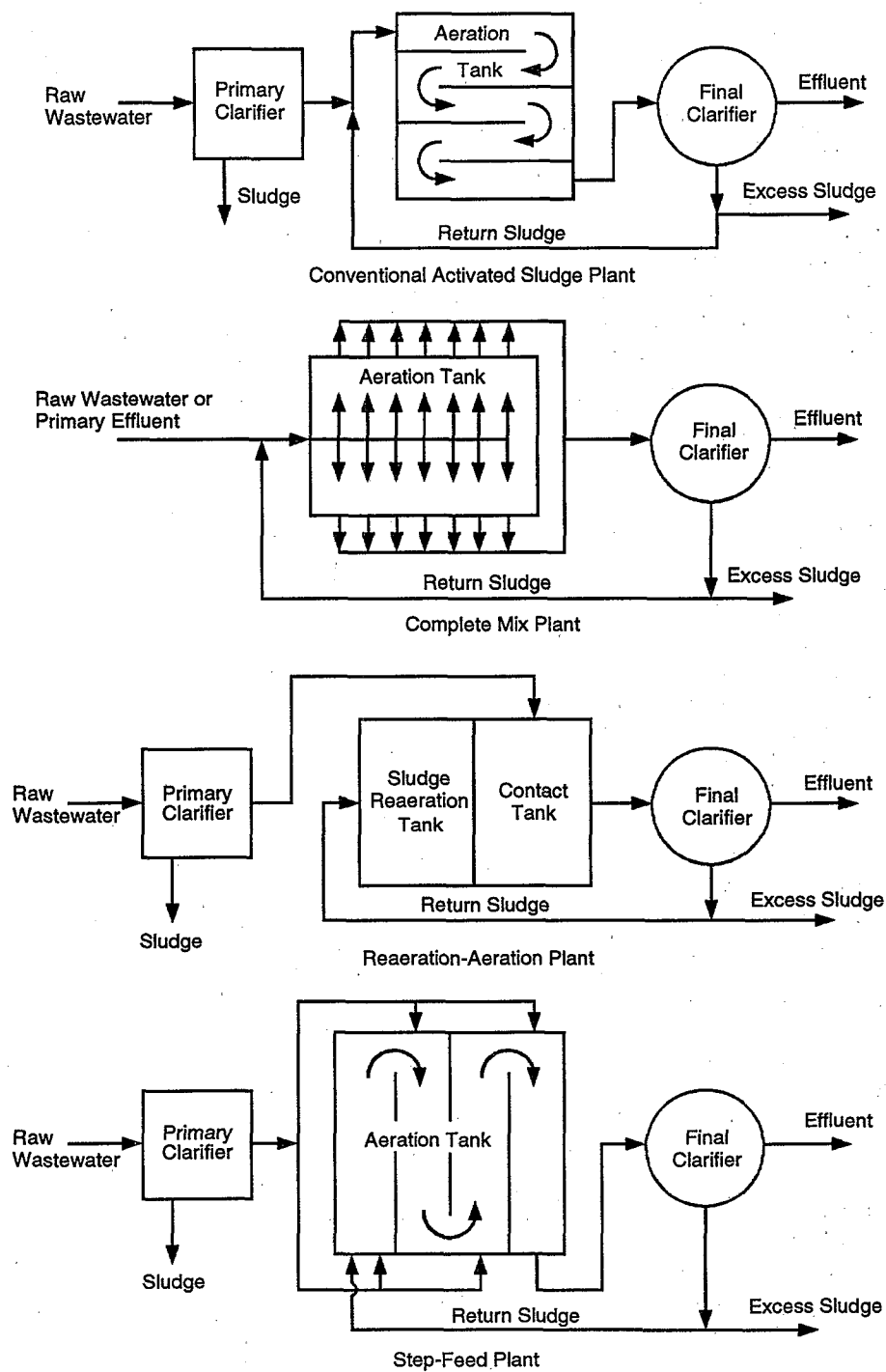
Figure 6-2. Suspended growth reactor configurations.

pertaining to each suspended growth configuration are presented in Sections 6.4.2 through 6.4.9. In Section 6.4.1, the nitrification kinetic theory presented in Chapter 3 is applied to the design of suspended growth reactors.

6.4.1 Application of Kinetic Theory to Design

The nitrification kinetic theory presented in Chapter 3 may be directly applied to the design of suspended growth reactors that are compatible with nitrification. The equations must be adapted to the specific configuration under consideration, but in most cases this adaptation is relatively straightforward.

Nitrification kinetic theory can be applied to define the following parameters:

1. The peaking factor required to handle diurnal transients in loading to prevent significant ammonium bleedthrough under peak load conditions.

2. The minimum and design solids retention times under the anticipated conditions of pH, DO, and temperature. As discussed in Section 6.4.10.1, the design solids retention time is calculated through the use of an overall process design factor, which accounts for influent loading variations and other factors, such as process variability (e.g., DO), uncertainty in kinetics, and the potential presence of inhibitory compounds.

3. The observed organic removal rate and the observed ammonium oxidation rate specific to the feed C to N ratio, based on measurement of the total reactor VSS.

4. The required hydraulic retention time in the suspended growth reactor and the reactor volume.

5. The excess sludge wasting schedule.

Determining the design solids retention time is the first step in sizing the suspended growth nitrification reactor. It follows from the relationship between the net specific growth rate (μ'_N) and the solids retention time (θ_c) of the organisms in the reactor that has been presented in Chapter 3:

$$\theta_c = \frac{1}{\mu'_N} \qquad (3\text{-}13)$$

and recognition of the need to ensure that Equation 3-17 is satisfied:

$$\theta_c^d \geq \theta_c^m \qquad (3\text{-}17)$$

Thus, the solids retention time approach to sizing nitrification reactors has a fundamental basis. The procedure classically has been simplified by making certain assumptions when specifying the observed or net yield (Y_{NET}) of total VSS in the suspended growth system. As discussed in Section 3.3.6, and further detailed in Chapter 5, the levels of feed inert VSS, organism decay and other factors will influence the value of the observed yield. Chapter

5 presents information which allows application of the solids retention time design approach, but in a more rigorous fashion, accounting for many of these assumptions.

The use of experimentally measured ammonium oxidation or nitrification rates for reactor sizing is also valid. But the limitations of this rate approach must be realized, as previously discussed in Section 3.3.6.

6.4.1.1 Solids Retention Time Design Approach

The first step in sizing the suspended growth nitrification reactor is the selection of the solids retention time. To do this, one must account for the effects of the various kinetic factors on the growth rate of *Nitrosomonas*. The combined kinetic expression for *Nitrosomonas* growth, accounting for the effects of ammonium-nitrogen concentration and temperature according to Equations 3-10 and 3-14, is:

$$\mu_N = 0.47[e^{0.098(T-15)}]\left[\frac{N}{K_N + N}\right] \qquad (6\text{-}1)$$

This applies under the constraints that the reactor design ensures, nonlimiting DO and pH conditions according to the recommendations of Sections 3.3.3 and 3.3.4. For example, at T = 20°C, N = 2.5 mg/L and under nonlimiting DO and pH conditions, Equation 6-1 would yield:

$$\mu_N = 0.47 \ (1.63) \ (0.71) = 0.54 \ d^{-1}$$

In calculating μ_N, K_N was set at 1.0 mg/L, as suggested in Section 3.3.2. The calculated μ_N represents the maximum possible nitrifier growth rate under the environmental conditions specified at an NH_4^+-N concentration of 2.5 mg/L. The solids retention time corresponding to μ_N is the theoretical minimum solids retention time (θ_c^m) required under the specified environmental conditions and is calculated from Equation 3-13 with $b_N = 0$. That is:

$$\theta_c^m = \frac{1}{\mu'_N} = 1.84 \ d$$

Lawrence and McCarty introduced the concept of a safety factor (SF) in the application of biological treatment process kinetics to design (23). They noted that the safety factor, which was defined as the ratio of the design solids retention time to the minimum solids retention time, was necessary to ensure high treatment performance and process stability; to provide resistance to toxic upsets; and to minimize process variations caused by pH extremes, low DO concentration and inhibitory materials. As noted earlier in Section 6.4.1, design considerations also include consideration of a peaking factor (PF) to ensure that ammonium breakthrough does not occur during diurnal peaks in load. Both the SF and PF concepts are considered in selecting the overall design factor (DF). In some cases the PF and DF are equivalent. However, in other circum-

134

stances the product of the SF and PF may be more appropriate to establish the DF. The design solids retention time, resulting from the use of an overall DF, will typically be 1.5 to 3.0 times greater than the minimum solids retention time. Considerations in selecting the peaking and design factors are discussed in Section 6.4.10.1.

Once θ_c^d has been determined, Equations 3-13 and 3-10 can be used to calculate, sequentially, the design nitrifier growth rate and the actual steady state ammonium content of the effluent. Equation 3-10 must be adapted to reflect the hydraulic conditions characterizing the suspended growth reactor (e.g., complete mix versus plug flow) to calculate effluent ammonia levels. This consideration is illustrated in the design examples contained in Sections 6.4.2.2 and 6.4.3.2.

The concentration of active heterotrophic biomass in the reactor is represented by the following expression:

$$X_1 = \frac{Y_H}{1 + b_H\theta_c} (S_0 - S_1) \frac{\theta_c}{t} \qquad (6\text{-}2)$$

where the term $Y_H/(1 + b_H\theta_c)$ is the observed yield coefficient, Y_{NET}, for active biomass.

This expression in conjunction with Equation 3-19 will give the following expression:

$$\frac{1}{\theta_c} = Y_{NET}\, q_H \qquad (6\text{-}3)$$

As shown in Table 2-13, the classical approach to net VSS production, X, (i.e., the summation of active biomass plus endogenous decay products plus inert material in the entering wastewater) is an equation of the same form as Equation 6-2. The expression is:

$$X = \frac{Y(\Delta\ \text{substrate})}{1 + b\theta_c} \frac{\theta_c}{t} \qquad (6\text{-}4)$$

and this equation leads to the counterpart of Equation 6-3, namely:

$$\frac{1}{\theta_c} = Y_{NET}q_{OBS} \qquad (6\text{-}5)$$

where:

Y_{NET} = g total VSS produced/g COD removed (feed total COD minus effluent SCOD)

q_{OBS} = g COD removed/g total VSS/d

The value for Y_{NET} in Equation 6-5 (i.e., based on total VSS) is dependent on θ_c^d and the nature of the wastewater being treated (i.e., raw, primary, or secondary wastewater). The values presented in Figure 2-8 may be used for design in most cases, provided the wastewater does not contain a large industrial component. If site specific data are available, they should be used to establish the yield coefficients. The required hydraulic retention time, t, can

then be calculated from the definition of q_{OBS}, as derived from an alteration of Equation 3-19:

$$q_{OBS} = \frac{S_0 - S_1}{X\, t} \qquad (6\text{-}6)$$

where:

S_0 = feed or influent total COD, mg/L

S_1 = effluent soluble COD, mg/L

One must define an allowable level of MLVSS or X in order to use Equation 6-6. The allowable level of mixed liquor total suspended solids (MLTSS), and therefore X, is influenced primarily by the efficiency of the solids-liquid separation step (Section 6.4.10.5).

The observed ammonium oxidation rate (r_N), although not pertinent to design by this procedure, can be calculated by modifying Equation 3-21 as follows:

$$r_N = \frac{N_0 - N_1}{X\, t} \qquad (6\text{-}7)$$

The excess sludge wasting requirements can be calculated from the definition of the solids retention time according to Equation 3-12. The specific design examples presented in Sections 6.4.2.2 and 6.4.3.2 will illustrate in detail the use of the solids retention time design approach.

6.4.1.2 Alternative Design Approach

An alternative to the solids retention time design approach relies on the use of an ammonium oxidation rate, which is the ammonium oxidized per unit time at a specific VSS level in the system (g NH_4^+-N oxidized per day per g MLVSS). The ammonium oxidation rate is equivalent to the maximum rate if the rate is zero-order with respect to ammonium concentration, as previously discussed in Chapter 3 (Section 3.3.6). That is:

$$q_N = \hat{q}_N \qquad (3\text{-}20)$$

Determining the maximum oxidation rate effectively defines the minimum solids retention time, θ_c^m, through the relationship:

$$\theta_c^m = \frac{1}{Y_N\, \hat{q}_N} \qquad (3\text{-}23)$$

Often designers that use the ammonium oxidation rate approach have simply selected a nitrification rate and used that value to determine t for an allowable level of total mixed liquor VSS. This, in effect, is equivalent to using the observed ammonium oxidation rate, r_N, as defined by Equation 6-7. This approach is acceptable when it is based on rate information that is derived specific to the wastewater in question; the observed maximum nitrification rate relative to the actual, or true, maximum rate will reflect the level of active Nitrosomonas as VSS (X_N) as a fraction of the total VSS (X) in the reactor. That is,

135

the observed maximum rate, $\hat{r}_N$, is related to true maximum rate, $\hat{q}_N$, by the following equation:

$$\hat{r}_N = f \hat{q}_N \qquad (6\text{-}8)$$

where:

f = fraction of active *Nitrosomonas* VSS in the reactor total VSS

It is clear from Equation 6-8 that the fraction of nitrifiers present in the reactor has a marked effect on the observed nitrification rates. As has been discussed in Chapter 3 (Section 3.3.6), the fraction of nitrifiers present in the mixed liquor will strongly depend on the wastewater feed C:N. The influence of this ratio on nitrifier fraction and nitrification rates was recognized as early as 1940 (24). The wide range in reported nitrification rates which is demonstrated by Figure 6-3, is likely caused in part by differences in the wastewater C:N.

The limitations in using the ammonium oxidation or nitrification rate approach to size the nitrification reactor are obvious from the preceding discussion. There is an advantage to using the rate approach only when information on the site-specific rate is available, ideally through pilot studies. The approach, which is basically the same as the solids retention time approach, can also be used as a check against the solids retention based sizing to verify that the rate information is reasonable.

Once an observed ammonium oxidation rate has been determined, a design rate value can be established from the maximum observed rate, accounting for the effects of the required effluent ammonia-N concentration and other

factors through the use of a design factor. The design rate can then be used to establish t for an allowable level of X. Fixing t and X establishes the organic removal rate that will be observed (Equation 6-6), which, in turn, fixes the operating θ_c for a given observed total VSS yield (Equation 6-5).

6.4.2 Complete Mix Systems

Many activated sludge systems are designed to operate on the complete mix principle. When designed in this fashion the system is often referred to as the complete mix activated sludge or CMAS process. An example of the feed and withdrawal arrangement for a complete mix plant may be found on Figure 6-2. The complete mix design provides uniformity of load to all points within the aeration tank, easing the problems of oxygen transfer that can be encountered at the head end of conventional plants, and making this configuration attractive for handling organic shock loads. The contents of the complete mix reactor are homogenous and thus the mixed liquor feed to the clarifier is identical to the mixed liquor throughout the reactor. Complete mix systems can be prone to "bleed through" or short circuiting of the feed to the effluent, particularly during peak flows.

6.4.2.1 Design Approach

The solids retention time design approach presented in Section 6.4.1.1 can be directly applied to the design of complete mix activated sludge systems, using Equation 3-10 (Section 3.3.1) to calculate the actual steady state ammonium content of the effluent.
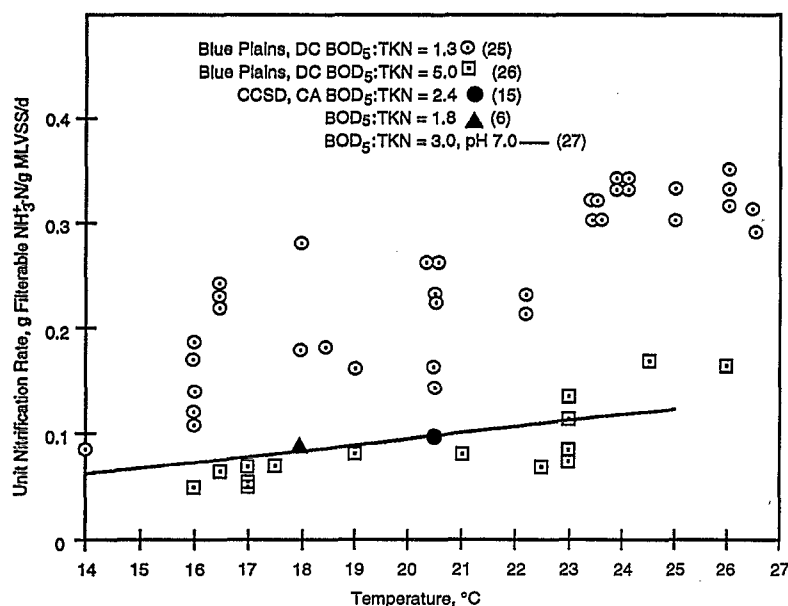
Figure 6-3. Variation in observed nitrification rates.

6.4.2.2 Design Example No. 1: Nitrification in a Complete Mix Suspended Growth System at High Carbonaceous Feed Concentration

The following design example presents the solids retention time design approach for the example "complex" Plant B. The calculations will address sizing requirements to meet the more stringent effluent limits. The design conditions and effluent limits have been presented in Chapter 2 (Section 2.9 and Table 2-10). Some of the design information from Chapter 2 is also summarized in Table 6-2. Design calculations are demonstrated at 15°C. The results at this temperature and at 10°C and 20°C are also presented for discussion purposes. A simplified process schematic is presented in Figure 6-4.

Table 6-2. Design Conditions for Example 1: Plant B in a Complete Mix Configuration with Higher Carbonaceous Feed and More Stringent Effluent Requirements

Wastewater Flow Characteristics, m³/d (mgd)		
Raw wastewater average flow	18,925 (5.0)	
Total influent average flow	21,055 (5.56)	
Primary Effluent Characteristics, mg/L		
COD	187[a]	168[b]
CBOD$_5$	97[a]	87[b]
TSS	80[a]	72[b]
TKN	29.5[a]	26.6[b]
Total P	6.0[a]	5.4[b]
Alkalinity, mg/L (as CaCO$_3$)		120[b]
Final Effluent Limits, mg/L		
CBOD$_5$	10	
TSS	10	
NH$_4^+$-N	2	
Total N	5	
Total P	1	
Design Conditions/Assumptions		
Reactor temperature, °C	15	
Reactor, MLVSS, mg/L	1,400	
Reactor minimum DO, mg/L	2.5	
Reactor pH range	7.0–7.6	
Reactor and effluent VSS/TSS	0.68[c]	

[a] Concentration values at average conditions expressed as mg/L equivalents (see Table 2-16).
[b] Concentration value at average conditions at total flow of 21,055 m³/d.
[c] Does not include impact of mineral addition for phosphorus removal.

Figure 6-4. Simplified schematic for Design Example No. 1 employing a complete mix suspended growth reactor for nitrification.

6.4.2.2 Design Example No. 1 (continued)

1. Determine the process design factor to be used in sizing the biological reactor. Following from Chapter 2 (Table 2-18), the peaking factor to be used in this example to account for influent loading variations is 1.56. Assume that the effluent quality requirement, the anticipated variations in process conditions, an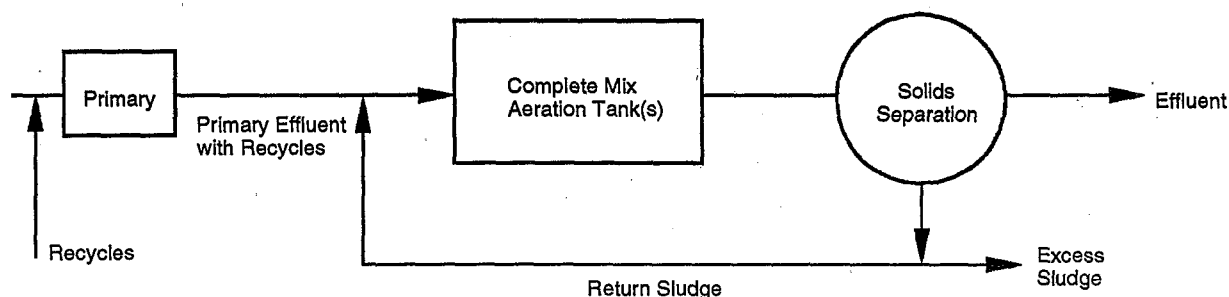d the uncertainty in the kinetic coefficients warrant a safety factor of 2.0. Compute an overall process design factor of 3.1 based on the product of PF and SF.

2. Verify that the reactor operating pH can be expected to be in the assumed range of 7.0 to 7.6 (Table 6-2). Alkalinity destroyed can be calculated from the amount of ammonia nitrogen assumed to be oxidized and from the coefficient presented in Table 3-1. From callout 6 in Table 2-16, the mg/L equivalent soluble nitrogen was 25.7 mg/L. Allowing for about 1 mg/L equivalent of soluble refractory nitrogen, the actual concentration of oxidizable nitrogen in the secondary effluent may be estimated as 24.7/1.09955 or 22.5 mg/L. Even though the effluent limit is 2 mg/L of NH_4^+-N, the system will normally produce much lower residual effluent NH_4^+-N concentrations. Therefore, assume the alkalinity destroyed is given by:

(7.1 mg $CaCO_3$/mg N oxidized) (22.5 mg/L) = 160 mg/L (as $CaCO_3$)

Alkalinity available (Table 6-2) is 120 mg/L (as $CaCO_3$). From Chapter 2 (Table 2-3), and assuming that an air oxygen transfer system with only 12-percent efficiency is used (see Step 11), it is reasonable to assume that the pH will remain in the required range (Section 6.4.10.3), provided a minimum residual alkalinity of 50 mg/L (as $CaCO_3$) is maintained. On this basis, the required supplemental alkalinity will be:

(160 mg/L destroyed + 50 mg/L minimum residual − 120 mg/L available)
= 90 mg/L (as $CaCO_3$) to be supplemented

The mass of alkalinity required under average day conditions will be:

$$\frac{21,055 \text{ m}^3/\text{d} \ (90 \text{ g/m}^3)}{10^3 \text{ g/kg}} = 1,895 \text{ kg} \ (4,180 \text{ lb})/\text{d} \ (\text{as } CaCO_3)$$

The design example peaking factors presented in Table 2-12 can be used to determine alkalinity requirements under other loading conditions. For example, the nitrogen mass increases by a factor of 1.7 on the maximum day with a corresponding increase in matching alkalinity of only 1.5. Influent flow increases by a factor of 2.5, and, for this example calculation, it is assumed that return flows to the head of the plant also increase by this factor. Thus the mass of alkalinity required on the maximum day can be estimated as follows:

Flow = 2.5 (21,055) = 52,638 m^3/d

Influent NH_4^+-N = (1.7/2.5) (22.5) = 15.3 mg/L

Influent alkalinity = (1.5/2.50)(120) = 72 mg/L as $CaCO_3$

Supplemental alkalinity = [(7.1)(15.3) − 72 + 50] [52,638/10^3] = 4,560 kg (10,050 lb)/d as $CaCO_3$

Additional alkalinity considerations where metal addition (Al^{+3} or Fe^{+3}) is used for phosphorus removal are presented in Step 12.

3. Calculate the maximum nitrifier growth rate under the nonlimiting DO and pH conditions stated (Sections 3.3.3 and 3.3.4). Equation 3-14 should be used to calculate $\hat{\mu}_N$ under nonlimiting substrate conditions (i.e., $K_N \ll N$):

$$\hat{\mu}_N = 0.47[e^{0.098(T-15)}] \qquad \qquad (6-9)$$

At T = 15°C (Table 6-2):

$$\hat{\mu}_N = 0.47 \text{ d}^{-1}$$

4. Calculate the minimum solids retention time for nitrification. From Equation 3-13 with $b_N = 0$, the correct expression is:

$$\theta_c^m = \frac{1}{\hat{\mu}_N} \tag{6-10}$$

For this example:

$$\theta_c^m = \frac{1}{0.47} = 2.13 \text{ d}$$

5. Calculate the design solids retention time according to:

$$\theta_c^d = (\text{Process design factor}) \, \theta_c^m \tag{6-11}$$

For this example:

$$\theta_c^d = 3.1 \, (2.13) = 6.6 \text{ d}$$

6. Calculate the design nitrifier growth rate. From Equation 3-13 with $b_N = 0$, the correct expression is:

$$\mu_N = \mu'_N = \frac{1}{\theta_c^d} \tag{6-12}$$

For this example:

$$\mu_N = \frac{1}{6.6} = 0.152 \text{ d}^{-1}$$

7. Calculate the steady state ammonium content of the effluent. Equation 3-10 is directly applicable to complete mix activated sludge systems, where N_1 is the effluent or reactor ammonium-nitrogen content:

$$\mu_N = \hat{\mu}_N \frac{N_1}{K_N + N_1} \tag{6-13}$$

where: N_1 = effluent NH_4^+-N, mg/L

Select a value of 1.0 for K_N, according to the recommendation in Section 3.3.2. Equation 6-10 becomes:

$$\mu_N = 0.152 = 0.47 \frac{N_1}{1.0 + N_1}$$

therefore: $N_1 = 0.48$ mg/L

8. Calculate the organic removal rate. Following from Chapter 2 (Figure 2-10), at θ_c^d of 6.6 days, Y_{NET} is approximately 0.25 g total VSS produced/g COD removed. Therefore, for this example:

$$q_{OBS} = \frac{1}{\theta_c^d \, Y_{NET}} \tag{6-14}$$

$$q_{OBS} = \frac{1}{(6.6)(0.25)} = 0.606 \text{ g COD/g MLVSS/d}$$

Where plant data are sufficient to develop a relationship between Y_{NET} and θ_c, these data should be used in place of Figure 2-10. If the municipal wastewater contains an influent composition where the ratios of degradable VSS to inert VSS are substantially different than used in the example, the net yield should be calculated for the wastewater in question (see Table 2-13). Finally observed yields are temperature sensitive, (28) and this additional factor can be considered when calculating net yields. Given the many uncertainties normally existing with regard to influent wastewater composition and flows, the simple procedure used in this example will prove adequate for many cases.

6.4.2.2 Design Example No. 1 (continued)

9. Determine the reactor hydraulic retention time and the reactor volume at total influent average flow. From Equation 6-6, the correct expression for t is:

$$t = \frac{S_o - S_1}{q_{OBS} X} \tag{6-15}$$

Using the values from Table 6-2 for S_o and X, and an estimate of 18 mg/L for S_1 (Table 2-16, callout 6) gives:

$$t = \frac{168 - 18}{(0.606)(1,400)} = 0.176 \, d = 4.2 \, hr$$

The reactor volume can be calculated from:

$$Volume = Q \, t \tag{6-16}$$

where: Q = total influent average flow rate, m^3/d

For this example:

$$Volume = (21,055)(0.176) = 3,710 \, m^3 \, (131,000 \, cu \, ft)$$

10. Determine the sludge wasting requirements. Sludge is wasted from the system as solids contained in the effluent from the secondary clarifier and intentionally from the reactor or the return sludge stream. The total sludge to be wasted under steady state conditions can be calculated from the definition of the solids retention time (Equation 3-12) expressed for a suspended growth reactor as:

$$\theta_c^d = \frac{I_A}{S} \tag{6-17}$$

where:

I_A = inventory of VSS under aeration, kg

S = total VSS wasted, kg/d

The inventory of VSS under aeration can be calculated according to:

$$I_A = \frac{XV}{10^3} \tag{6-18}$$

where: V = volume of aeration tank, m^3

For this example:

$$I_A = \frac{(1,400)(3,710)}{10^3} = 5,190 \, kg \, (11,440 \, lb) \, VSS$$

Using Equation 6-17 and θ_c^d of 6.6 days, the total VSS to be wasted is:

$$S = \frac{5,190}{6.6} = 786 \, kg \, (1,1730 \, lb) \, VSS/d$$

The sludge contained in the effluent will depend upon the efficiency of the secondary clarifier. For purposes of this example, assume that the effluent TSS are estimated as 14 mg/L and the VSS:TSS ratio in the reactor is 0.68 (without chemical addition for P removal). The VSS contained in the effluent is:

$$\frac{21,055(14)(0.68)}{10^3} = 200 \, kg \, VSS/d \, (442 \, lb/d)$$

6.4.2.2 Design Example No. 1 (continued)

Therefore, the sludge to be wasted from the mixed liquor or return sludge is:

$$786 - 200 = 586 \text{ kg (1,290 lb) VSS/d}$$

$$586/0.68 = 862 \text{ kg (1,900 lb) TSS/d}$$

Where there is supplemental chemical addition to the mixed liquor for phosphorus removal (see Step 12), the sludge wasting calculations and estimated VSS:TSS must also include the impact of the additional inert solids. Solids-liquid separation considerations (Section 6.4.10.5) will dictate clarifier design including the return sludge rate and concentration. The mass of sludge to be wasted can then be calculated for average and maximum month conditions.

11. Determine aeration requirements. Oxygen requirements for ammonium oxidation can be calculated from the coefficient presented in Table 3-1. Oxygen requirements for CBOD oxidation are dependent upon the degree of biomass oxidation as illustrated in Figure 2-10. The COD oxidized can be computed as the difference between the influent COD and the COD discharged from the system. The discharged COD is the sum of the soluble effluent COD and the COD of the VSS exiting the system. At the 6.6-day design θ_c, the COD in the effluent VSS can be estimated as:

$$(786 \text{ kg VSS/d})(1.42 \text{ kg COD/kg VSS}) = 1,120 \text{ kg (2,470 lb) COD/d}$$

Carbonaceous oxygen requirements are:

$$\frac{21,055(168 - 18)}{10^3} - 1,120 = 2,040 \text{ kg (4,500 lb) O}_2\text{/d}$$

An alternative "rule-of-thumb" methodology to estimate carbonaceous requirements is to select a coefficient value of from 1.0 to about 1.3 (for high θ_c systems) times the BOD$_5$ removal (influent minus soluble effluent). For a BOD$_5$ removal of 85 mg/L (Table 6-2 and assuming a soluble effluent BOD$_5$ of 2 mg/L) and a total carbonaceous requirement of 2,040 kg O$_2$/d, the equivalent coefficient value would be 1.14.

Oxygen required for ammonium oxidation (ignoring the small effluent NH_4^+-N residual) is:

$$\frac{21,055(22.5)(4.6)}{10^3} = 2,180 \text{ kg (4,806 lb) O}_2\text{/d}$$

Total oxygen required under average conditions is:

$$2,040 + 2,180 = 4,220 \text{ kg (9,300 lb) O}_2\text{/d}$$

$$\left(\frac{4,220}{3,710}\right)\left(\frac{10^3}{24}\right) = 47.4 \text{ mg O}_2\text{/hr/L of aeration tank}$$

The design example peaking factors presented in Chapter 2 (Table 2-12) can be used to determine oxygen requirements under other conditions. For example, under maximum day conditions the peak oxygen requirement will be approximately:

$$2,040(2.1) + 2,180(1.7) = 7,990 \text{ kg (17,610 lb) O}_2\text{/d}$$

$$(7,990/3,710)(10^3/24) = 89.7 \text{ mg O}_2\text{/hr/L of aeration tank}$$

The efficiency of the oxygen transfer system will dictate the amount of air required to satisfy the total oxygen demand. The oxygen transfer efficiency depends on a number of factors such as type of aeration device, reactor depth, alpha value, and operating DO level. An oxygen transfer efficiency of 10 percent requires

6.4.2.2 Design Example No. 1 (continued)

approximately 36.2 m³ of air/kg O_2 transferred (580 cu ft/lb O_2). For this example, assume an overall transfer efficiency of 12 percent. Therefore, the aeration capacity required at average conditions is:

$$36.2 \frac{(0.10)}{(0.12)} \frac{(4,220)}{(1,440)} = 88.4 \text{ m}^3/\text{min (3,120 scfm)}$$

The aertion requirement for the maximum day will be:

$$\frac{7,990(36.2)(0.10)}{(1,440)(0.12)} = 167.4 \text{ m}^3/\text{min (5,910 scfm)}$$

More detailed information on aeration requirements and design for suspended growth reactors can be found in Section 6.4.10.3.

12. Determine impact of phosphorus removal on design. When phosphorus removal is required, the design must address whether a system with anaerobic and aerobic zones would be appropriate. Phosphorus removal can be obtained by relying on chemical addition or by using chemical addition as needed to supplement an enhanced biological removal process. Effluent filtration may also be required.

Chemical addition affects the design approach with regard to alkalinity requirements (Step 2) and the production of additional SS. The additional SS can affect aeration basin and clarifier sizing and will increase waste solids production and sludge handling requirements. Information specific to this issue is provided in Sections 6.4.10.2 and 2.7.2 and also Table 2.3 as well as in other references (2,29,30).

For the example problem, the influent mass of phosphorus is 113.6 kg/d and approximately 15.7 kg/d [786 kg VSS x 0.02] is associated with the waste VSS. Addition of $FeCl_3$ or $Al_2(SO_4)_3 \cdot 14 \, H_2O$ could be selected for the required P removal.

In calculating additional solids production and alkalinity losses by metal addition, there are two major pH-dependent competing reactions as illustrated by the following reactions for aluminum.

$$1) \quad Al^{+3} + HCO_3^- \Rightarrow Al(OH)_3 + 3 \, CO_2$$

$$2) \quad Al^{+3} + PO^{-3} \Rightarrow AlPO_4$$
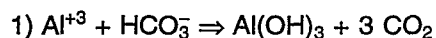
For the first reaction, there are 5.6 mg $CaCO_3$ alkalinity lost per mg Al reacting and 2.9 mg of $Al(OH)_3$ formed per mg Al reacting. For the second reaction, there is 3.9 mg $AlPO_4$ formed per mg P removed. To achieve the effluent limit of 1 mg/L total P will probably require from 1.25 to 1.75 moles of Al per mole of P. This chemical addition could be split among the primary clarifier, the biological reactors (see Step 16) and the effluent filters.

An estimate of solids production at 1.5 moles Al/mole P follows:

From 1:   (1.5 − 1.0)(27/31)(2.9)(113.6 − 15.7) = 124 kg (273 lb) $Al(OH)_3$/d

From 2:   3.9 (113.6 − 15.7) = 382 kg (842 lb) $AlPO_4$/d

Total inert solids production equals 506 kg additional TSS/d. Alternatively, one can more easily estimate the additional inert solids production from the coefficient in Table 2.3 as follows:

   (6.1)(113.6 − 15.7) = 597 kg (1,320 lb) TSS/d

The coefficient 6.1 assumes somewhat different stoichiometry than represented by Reactions 1 and 2 above, but is an equally acceptable approach for estimation purposes.

6.4.2.2 Design Example No. 1 (continued)

Since θ_c^d is 6.6 days, the reactor will carry an additional $(6.6)(506) = 3,340$ kg (7,365 lb) inert TSS or an additional 900 mg/L of additional TSS in the aeration tank. The total MLSS concentration will be approximately

$$\frac{1,400}{0.68} + 900 = 2,960 \text{ mg TSS/L}$$

This solids estimate should be used to revise the waste sludge production previously presented in Step 10 (which did not consider the impact of alum addition).

The additional alkalinity depletion from the alum addition is:

$(0.5)(27/31)(113.6 - 15.7)(5.6) = 239$ kg (527 lb) as $CaCO_3$/d

This additional alkalinity loss was not included in Step 2 and would need to be considered when computing supplemental alkalinity requirements.

13. Evaluate at alternative temperatures (Table 6-3).

Table 6-3. Summary of Example 1 Design Results: Plant B at Higher Carbonaceous Feed Concentration and More Stringent Effluent Requirements

Design Parameter	Value as a Function of Temperature		
	10°C	15°C	20°C
$\hat{\mu}_N$, d^{-1}	0.288	0.47	0.767
θ_c^m, d	3.47	2.13	1.30
θ_c^d, d	10.8	6.6	4.0
μ_N, d^{-1}	0.093	0.152	0.25
N_1, mg/L	0.48	0.48	0.48
Y_{NET}, g VSS/g COD	0.22	0.25	0.30
t, hr	6.1	4.2	3.1
Reactor volume, m^3	5,630	3,710	2,710
VSS inventory (I_A), kg	7,500	5,190	3,790
VSS wasted (S), kg/d	694	786	948
Average carbonaceous O_2 demand, kg/d	2,170	2,040	1,810
Average total O_2 demand, kg/d	4,350	4,220	3,990
Average O_2 demand, mg/L of aeration tank/hr	33.8	47.4	61.3

14. Consider evaluating the chosen design with one of the dynamic models discussed in Chapter 5 to investigate the actual dynamic response in ammonium nitrogen levels under the plant diurnal loading conditions.

15. Consider the design calculations as just one component is sizing the reactor system. Other factors, as discussed in Chapter 2, are equally important. In some instances, especially for small facilities, prudent design anticipates that administrative and/or operator support may not be sufficient to ensure that the design assumptions are fully realized. These constraints can be partially or fully accommodated with larger systems than dictated solely by kinetic considerations. Thus for the design example, an SRT of from 10 to 15 days could also be more appropriate when other intangible factors affecting plant operation are also evaluated.

16. To achieve the final effluent limit of 5 mg/L TKN specified in Table 6-2, this design approach requires the addition of a separate stage denitrification system to reduce the NO_3^--N. Design of suitable separate stage systems is discussed in detail in Chapter 7.

6.4.3 Conventional or Plug Flow Systems

Conventional systems consist of a series of rectangular tanks or passes (Figure 6-2) in which the ratio of total tank length to width is typically between 5 and 10. The hydraulics of this configuration have been loosely termed as plug flow because the influent wastewater and return activated sludge are returned to the head end of the process and the combined flow must pass along a long narrow aeration tank prior to exiting the system. The degree to which the process reactors actually approach plug flow is dependent on the amount of longitudinal mixing. While complete mix activated sludge systems are typical for a wide range of industrial application, plug flow hydraulic conditions are favored in the design of municipal wastewater nitrification reactors. As will be shown in Section 6.4.3.1, plug flow conditions result in a lower effluent ammonium concentration than a complete mix condition at the same design factor, or alternately, the same ammonium level at a lower design factor.

High-purity oxygen and powdered activated carbon nitrification reactors can be designed as two or more completely mixed tanks in series or as single tanks with a number of passes, thus incorporating hydraulic conditions equivalent to conventional plug flow activated sludge systems.

6.4.3.1 Design Approach

The approach to designing conventional activated sludge systems is similar to that for complete mix systems except for the equations used to predict effluent quality. A plug flow model may be applied to approximate the hydraulic regime in these systems, and the Monod expression for *Nitrosomonas* growth rate (Equation 3-10) must be integrated over the period of time an element of liquid remains in the nitrification reactor. The following is a solution for plug flow kinetics that can be adapted to this problem for conditions in which the sludge recycle ratio is less than 1.0 (23):

$$\frac{1}{\theta_c^d} + \frac{\hat{\mu}_N(N_o - N_1)}{(N_o - N_1) + K_N \ln\left[\dfrac{N_o}{N_1}\right]} \quad (6\text{-}19)$$

Reference 23 should be consulted to determine the form of Equation 6-19 for recycle ratios greater than 1.0.

Equation 6-19 is evaluated in Figure 6-5 for Design Example No. 1, presented in Section 6.4.2.2. This is shown as a function of the process design factor, DF, recognizing that from Equation 6-11:

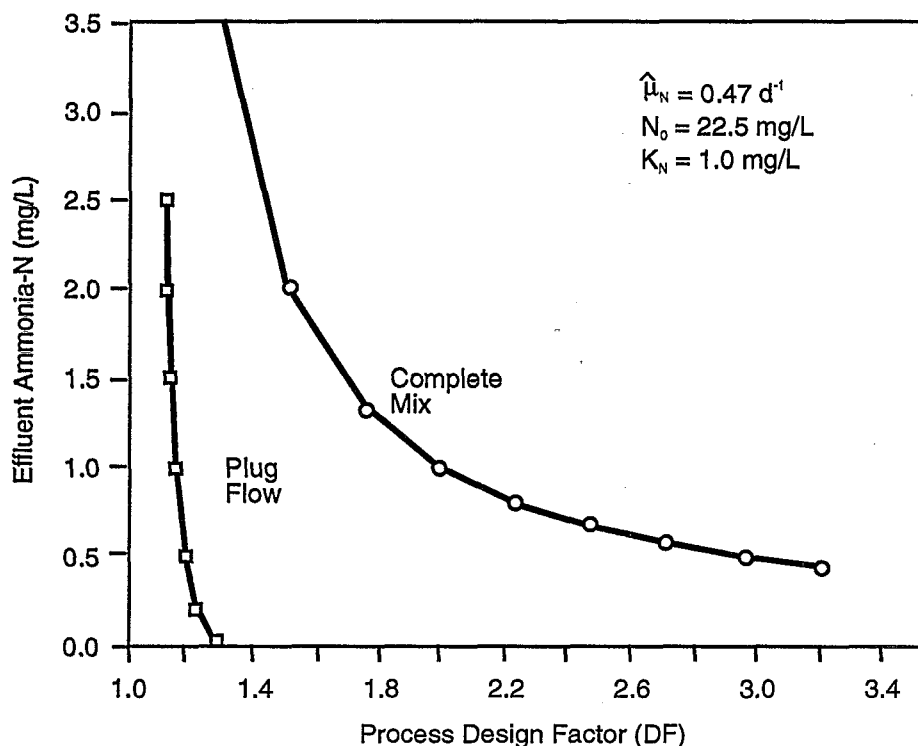$$DF = \frac{\theta_c^d}{\theta_c^m} \quad (6\text{-}20)$$

Figure 6-5. Effect of design factor on steady state effluent ammonia levels in complete mix and plug flow suspended growth reactors.

Lower values of the process design factor can be used for plug flow nitrification reactors to produce the same theoretical effluent ammonium concentration found in a complete mix reactor (Figure 6-5). This means that plug flow processes are theoretically more efficient at the same process design factor, or alternatively, require less aeration tank volume for the same level of nitrification efficiency.

However, plug flow type reactors also have a potential disadvantage in that the carbonaceous oxygen demand is concentrated at the head end of the tank, sometimes making it difficult to supply enough air in that area for both carbonaceous oxidation and nitrification. Air diffusion systems must be specifically designed to handle this concentrated load in higher feed $CBOD_5$ systems; otherwise, the first portion of the tank will not be available for nitrification because of inadequate levels of DO. A typical DO and nitrification pattern for plug flow tanks in which aeration capability is limited in the front end of the tank is presented in Figure 6-6. As can be seen from the aeration tank DO and ammonium nitrogen profile, nitrification is inhibited in the first portion of the tank because of the DO suppression that has been incurred for carbonaceous oxidation. Once the DO rises, the ammonium level falls at a reaction rate that approximates zero order, a reaction order predicted by kinetic theory (Section 3.3.1). It is notable that if sufficient aeration capability had been available in the head end of the tank, virtually complete nitrification probably would have been obtained. If oxygen

supply limitations are present in the head end of the tank, the plug flow type reactor's advantage over the complete mix reactor is reduced. The tapered aeration process modification to reactors designed with plug flow hydraulics represents a solution to this problem (Section 6.4.6).

The degree to which nitrification reactors approach plug-flow operation can be examined through dispersion theory (32,33). Reactors can be characterized by a dimensionless axial dispersion number, D/uL, where D is the axial dispersion coefficient in m^2/hr, u is the mean displacement velocity along the tank length in m/hr and L is the tank length, m. In the calculation of the axial dispersion number, u and L are known for any particular plant design and D must be measured. An acceptable approximation of D for both fine and coarse bubble diffused air plants can be obtained from the following relationship (33).

$$D = 3.118 \ W^2 \ (A)^{0.35} \qquad (6-21)$$

where:

W = tank width, m
A = air flow per unit tank volume, in standard $m^3/min/1,000 \ m^3$

The axial dispersion coefficient, D, is zero for true plug flow plants and infinite (∞) for true complete mix plants. Plants with a dispersion number, D/uL, less than 0.2 are usually classified as plug flow reactors, while for complete
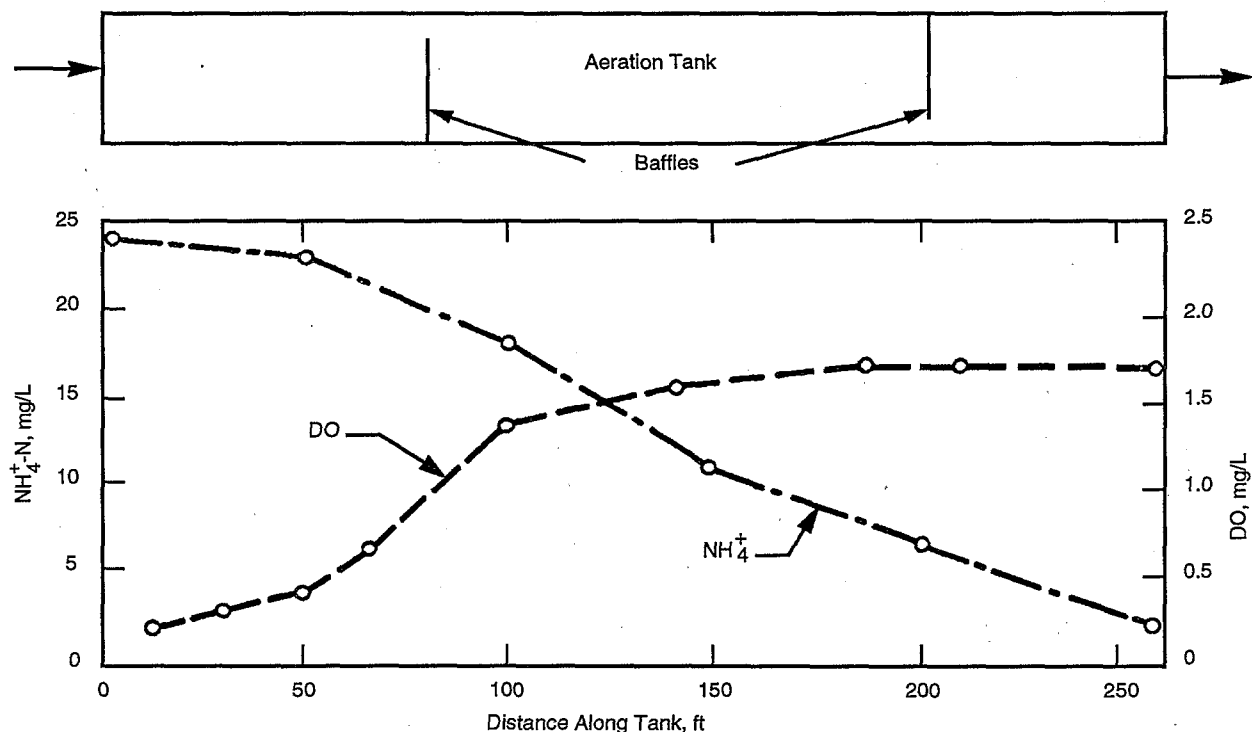

Figure 6-6. DO and ammonium-nitrogen profile in a plug flow system (from Reference 31).

145

mix systems, D/uL is typically greater than 4.0 (34). As an example calculation, assume a system's four-pass nitrification tanks have the following characteristics:

Air flow = 51.1 standard m³/min/1,000 m³

Width = 10.7 m (35.1 ft)

Depth = 4.5 m (14.8 ft)

Cross-sectional area of tank = 48 m² (525 sq ft)

Total length = 329 m (1,080 ft)

Flow each tank (four passes) @ 50% recycle = 85,200 m³/d (22.5 mgd)

From these data, the mean displacement velocity is calculated to be 74.0 m/hr (243 ft/hr). From Equation 6-21, the dispersion coefficient is:

$$D = 3.12(10.7)^2(51.1)^{0.35} = 1,415 \text{ m}^2/\text{hr} \ (15,234 \text{ sq ft/hr})$$

and

$$D/uL = 0.058$$

At 0.058, D/uL is significantly less than 0.2; thus, the system's nitrification tanks closely approach plug flow hydraulics. Equation 6-21 can be utilized to evaluate mixing in actual plant designs. If they closely approach true plug flow, Equation 6-19 can be used to describe nitrification. It is probable that most plants operated in the conventional mode do approach plug flow. For those plants with intermediate values of D/uL, complete-mix kinetics can be employed, which will yield conservative answers.

The hydraulic configuration of nitrification tanks can also be designed to discourage back mixing by employing a series of complete mix tanks. Absolute prevention of back mixing can be achieved by the use of mixed liquor overflow weirs between reactors.

6.4.3.2 Design Example No. 2: Plant B—Separate Stage Nitrification at Low Carbonaceous Feed Concentration

This design example is based on information provided in Section 2.9 for the more complex Plant B. In this example it is assumed that the environmental conditions and the solids retention time in the existing biological reactor are such that only CBOD removal is accomplished. Consequently, a second stage biological reactor is required for nitrification of the secondary effluent. The design information utilized is summarized in Table 6-4.

Table 6-4. Design Information for Nitrification of a Low Carbonaceous Feed Concentration

Wastewater Flow Characteristics, m³/d (mgd)		
Raw wastewater average flow	18,925 (5.0)	
Total secondary effluent average flow	21,055 (5.5)	
Unsettled Secondary Effluent Concentration[a]		
Total COD, mg/L	111[b]	101[c]
Nitrogen available for nitrification, mg/L	23.1[b]	21[c]
Soluble COD mg/L	30[b]	27[c]
Alkalinity, mg/L (as CaCO₃)		120[c]
Nitrification Reactor Effluent Characteristics		
Soluble, COD, mg/L	20[b]	18[c]
VSS, mg/L	9[b]	8[c]
Ammonium nitrogen, mg/L	2.0[d]	
Design Conditions/Assumptions		
Reactor temperature, °C	15	
Reactor MLVSS, mg/L	1,400	
Reactor minimum DO, mg/L	2.5	
Reactor pH range	7.0–7.6	

[a] See Processing Point 4, Table 2-16.
[b] Concentration values at average conditions expressed as mg/L equivalents.
[c] Concentration values at the total flow of 21,055 m³.
[d] For effluent objective, see Table 2-10.

6.4.3.2 Design Example No. 2 (continued)

The design procedure is similar to that used for treating a high carbonaceous feed in a complete mix suspended growth reactor (Section 6.4.2.2). Plug flow hydraulic conditions in the nitrification reactor and the level of pretreatment (i.e., primary clarification and a first stage biological reactor) may warrant selection of an SF lower in value than that used in the complete mix reactor design example. Also the peak load may be attenuated in the first stage biological reactor depending upon reactor hydraulic characteristics (i.e., degree of mixing and hydraulic retention time, HRT). If an SF of 1.5 is selected, and the process design factor is computed based on the product of PF and SF, the design solids retention time becomes:

$$\theta_c^d = DF \cdot \theta_c^m \qquad (6\text{-}11)$$

$$= 1.56(1.5)(2.13) = 4.98 \text{ d}$$

The corresponding μ'_N calculated from Equation 3-13 is 0.20 d^{-1}. The steady state ammonium content for the plug flow reactor conditions can be calculated from Equation 6-19 to be less that 0.1 mg/L. For a complete mix reactor, effluent NH_4^+-N would be 0.74 mg/L. Calculating the organic removal rate, hydraulic retention time and reactor volume require specification of the observed yield, Y_{NET}. The fact that the CBOD or COD from the first stage reactor is largely due to biological solids makes it more difficult to estimate Y_{NET}.

The VSS concentration in the second stage nitrification system will be largely influenced by the efficiency of the secondary clarifier (unless some primary effluent is directly bypassed to the nitrification second stage or unless some other source of VSS such as waste secondary sludge is added to the nitrification reactor).

Assume that the secondary system operates at an SRT of 2 days and the residual effluent degradable SCOD which will be removed in the nitrification reactor at the 5-day θ_c^d is approximately 10 mg/L. The MLVSS in the nitrification reactor include the following components:

- Inert VSS which are in the entering wastewater including inert particulate biomass decay products produced in the secondary system.

- Active heterotrophic biomass entering the system plus a small amount grown on the entering biodegradable COD.

- Inert decay products produced in the nitrification reactor from the entering and synthesized active biomass.

- Autotrophic microorganisms.

If all entering nitrogen available for nitrification (21 mg/L) was converted to NO_3^--N, the net autotrophic yield would only be 2.1 mg of nitrifier VSS (Table 3.1) per L of entering wastewater. Ten mg/L of SCOD removal at a 5-day θ_c in the nitrification reactor would produce an additional heterotrophic net yield, Y_{NET}, of only 2.5 milligrams per liter of entering wastewater (Table 2-13, IAWPRC model). The sum of the new heterotrophic and nitrifier VSS from biomass growth, 4.6 mg/L, is less than the projected 8 mg/L of effluent VSS from the nitrification system. Hence, the solids balance in the nitrification reactor will be largely controlled by the amount of VSS entering from the secondary clarifier (or added from other sources as described in Section 6.3).

Assume that the secondary effluent VSS are 20 mg/L. A simple way to estimate change of this material in the nitrification reactor is to use the classical approach in Table 2-13. For each liter of entering wastewater and at the 5-day θ_c^d the 20 mg/L of entering VSS will be reduced to:

$$Y_{NET} = \frac{20}{1 + 0.06(5)} = 15.4 \text{ mg/L}$$

Therefore, the total solids balance accounting for new growth and biomass decay per liter of entering wastewater is:

$$X = 4.6 + 15.4 = 20 \text{ mg/L VSS}$$

It is also possible to use the IAWPRC approach in Table 2-13 to estimate the change in the entering VSS components. Whether or not this extra effort is justified depends on the level of knowledge of the various VSS components in the secondary system and the degree of confidence in the estimate for the secondary effluent VSS levels.

The net VSS production is 20 mg/L of entering wastewater, the design MLVSS level is 1,400 mg/L and the θ_c^d is 5 days; the required hydraulic retention time can be computed based on the definition of SRT (i.e., θ_c = mass of solids under aeration divided by the mass of solids wasted per day) as follows:

$$1,400 = \frac{20(5)}{t} \qquad t = 0.071 \text{ d} = 1.71 \text{ hr}$$

Reactor volume = 21,055(0.071) = 1,504 m³ (53,106 cu ft)

I_A = 1,400 (1,504)/10³ = 2,106 kg (4,640 lb) VSS

S = I_A/θ_c^d = 2,106/5 = 421 kg (930 lb) VSS/d

Effluent VSS = 21,055(8)/10³ = 168 kg (370 lb) VSS/d

Sludge to be wasted = 421 − 168 = 253 kg (560 lb) VSS/d

Determine the aeration requirements:

Oxygen required for ammonia oxidation:

(21.0)(4.6)(21,055/10³) = 2,034 kg (4,480 lb) O₂/d

Assuming 1.42 kg COD/kg VSS, carbonaceous oxygen requirements can be estimated as:

(10 − 2.5(1.42) + 1.42(20 − 15.4))(21,055/10³) = 273 kg (602 lb) O₂/d

Total oxygen requirements average 2,307 kg (5,087 lb) O₂/d. This is a mean oxygen requirement of 64 mg/hr/L of aeration tankage. Given that the PF for the maximum monthly load is 1.56, the oxygen demand for the plug flow reactor will increase to 100 mg/hr/L of aeration tankage under these conditions. As in Case 1, the designer may elect to increase the aeration tank volume to reduce volumetric oxygen demands to a level compatible with the desired aeration equipment.

6.4.3.3 Performance Information

River Oaks, Hillsborough County, Florida

The River Oaks Advanced Wastewater Treatment Plant is located in Hillsborough County, Florida, discharging ultimately to Tampa Bay. In the mid-1980s, the plant's capacity was increased from 132 L/s (3 mgd) to 440 L/S (10 mgd) and the plant was upgraded to achieve total nitrogen removal (35). The configuration of the plant's aeration tanks results in a conventional plug flow system, as opposed to complete mix. A high carbonaceous feed concentration is imposed at the head end of the aeration tanks, following primary treatment. Carbonaceous oxidation and nitrification are accomplished in three separate, fine pore, diffused aeration tanks operating in series. A separate sludge reaeration tank is included as part of the system and sodium aluminate is added prior to clarifica-tion of the nitrified reactor contents to accomplish phosphorus removal. In addition to handling 440 L/s (10 mgd) of primary treated wastewater, the nitrification system is designed to handle filter backwash and recycle water from the downstream unit operations and return flow from the equalization basin, installed to accommodate excessive primary effluent during wet weather conditions. Relevant design information for the River Oaks treatment plant is summarized in Table 6-5.

The nitrification portion of the system must achieve less than 0.5 mg/L NH₄-N for the plant to meet its final effluent total nitrogen limits. Operating and performance information pertaining to the nitrification system over a 12-month period from August 1988 to July 1989 is presented in Tables 6-6 and 6-7, respectively. The excellent nitrification performance of this Florida plant is not surprising in light of the very favorable year-round climatic conditions, and

Table 6-5. River Oaks Advanced Wastewater Treatment Plant—Design Information for Carbonaceous Oxidation/Nitrification System Following Primary Clarification (Adapted from Reference 35)

Total Design Flow	50,000 m³/d (13.2 mgd)
Average influent flow	37,800 m³/d (10 mgd)
Filter backwash and recycle water	4,500 m³/d (1.2 mgd)
Return flow from equalization	7,600 m³/d (2 mgd)
Carbonaceous Oxidation/Nitrification Reactors (Secondary Treatment)[a]	
Number of aeration tanks	3
Number of reaeration tanks	1
MLSS, mg/L	3,270
MLVSS, mg/L	1,750
t, hr	3.7
θ, d	5.0
F/M, kg BOD$_5$/kg MLVSS/d	0.3
Secondary Sedimentation Tanks[b]	
Number	3
Hydraulic loading rate, m³/m²/d (gpd/sq ft)	25 (620)
Solids loading rate, kg/m²/d (lb/sq ft/d)	122 (25)

[a] Phosphorus removal accomplished by sodium aluminate addition to the nitrification system. F/M represents food-to-mass ratio. Stated values were calculated at total design flow.

[b] Hydraulic and solids loading rates to secondary sedimentation tanks based on total design flow.

Table 6-6. River Oaks Advanced Wastewater Treatment Plant—Carbonaceous Oxidation/Nitrification System Operating Conditions August 1988 to July 1989 (Adapted from Reference 35)

Parameter	Nitrification System Value* Average	Range
MLVSS, mg/L	2,550	1,600–3,350
MLTSS, mg/L	4,350	2,750–5,650
θ, d	16	8–25
F/M, kg BOD$_5$/kg MLVSS/d	0.12	0.09–0.15
Temperature, °C	27	24–30
Clarifier loadings		
Hydraulic, m³/m²/d (gpd/sq ft)	14.8 (364)	13.4–18.1 (330–445)
Solids, kg/m²/d (lb/sq ft/d)	125 (25.7)	76–198 (15.5–40.6)

* Monthly values.

low loading conditions experienced during this period. The average monthly temperature did not fall below 24°C (75°F), and the average solids residence time was 16 days.

Jackson, Michigan

In Jackson, Michigan, winter climatic conditions are less favorable for nitrification. The Jackson municipal wastewater treatment plant is a conventional activated sludge system consisting of seven two-pass, fine pore, diffused aeration tanks following primary clarification. Phosphorus removal is accomplished by ferrous chloride addition to the aeration tanks. The performance of this plant for a period in 1973 and 1974 was reported previously (1). At that time, coarse bubble diffusers were used for aeration. In 1989, the plant installed fine pore aeration and made other plant modifications (36). The maximum monthly average capacity of the current system is 10,160 kg BOD$_5$/d (22,400 lb/d). The plant typically processes 616 to 748 L/s (14 to 17 mgd) of raw wastewater.

Operating and performance results from 1990 are presented for the Jackson plant in Table 6-8. The DO in the carbonaceous oxidation/nitrification reactors typically exceeded 3.0 mg/L, and the pH was in the neutral range. During the coldest months the temperature was estimated at 10°C. The 1990 performance results indicate a deterioration in ammonia oxidation during January and February. Plant operating personnel said this deterioration was due to the presence of cyanide, nickel, or zinc in the wastewater rather than to temperature. The final effluent zinc concentration during January and February averaged 0.14 mg/L versus the yearly average of 0.09 mg/L. Operations at an electroplating shop occasionally contribute up to 15 percent of the influent to the plant and may be the source of these materials. As discussed in Section 3.3.5, such materials have been identified as potential nitrification inhibitors.

The results from the coldest months of 1991 appear to support the claim that lower temperatures are not affecting nitrification. The effluent ammonium-nitrogen values during January, February, and March of 1991 averaged 0.17 mg/L and the average sludge residence time was 11.7 days. It is worthwhile noting that in order to achieve an effluent NH_4^+-N of 1.0 mg/L or less at 10°C (50°F) in a complete mix system, nitrification theory (Equation 6-1) would predict that a minimum solids retention time of 7 days would be required (assuming K_N is 1.0 mg/L and operations are under nonlimiting DO and pH conditions). The Jackson plant results appear to support this, although there are some inconsistencies in the reported MLVSS, θ, and t values (Table 6-8).

6.4.4 Extended Aeration and Oxidation Ditch Systems

Most extended aeration systems are similar to conventional and complete mix plants except that the hydraulic retention times range from 24 to 48 hours, instead of a retention time of less than 10 hours that is normally used in the conventional and complete mix configurations. As such, the system is often considered simply a process

149

Table 6-7. River Oaks Advanced Wastewater Treatment Plant—Carbonaceous Oxidation/Nitrification System Performance, August 1988 to July 1989 (Adapted from Reference 35)

Parameter*	Raw Wastewater Influent		Nitrified Effluent	
	Average	Maximum Month	Average	Maximum Month
Flow, L/S (mgd)	—	—	339 (7.7)	418 (9.5)
BOD_5, mg/L	142	176	4	8
TSS, mg/L	118	149	9	22
TN, mg/L	30.1	37.8	17.4	21.4
NH_4^+-N, mg/L	21.1	24.9	0.15	0.31
NO_2^--N plus NO_3^--N, mg/L	0.17	0.24	15.8	19.1
Org-N, mg/L	8.8	12.8	1.48	1.98
TP, mg/L	6.3	7.0	0.53	0.86

* TN represents total nitrogen; Org-N, organic nitrogen; and TP, total phosphorus.

Table 6-8. Jackson, Michigan, Wastewater Treatment Plant Nitrification System Operation Conditions and Performance Results, 1990 (Reference 36)*

Month	Wastewater Flow, L/s (mgd)	Solids Recycle Ratio	MLVSS, mg/L	θ, d	t, hr	BOD_5, mg/L		Ammonia-N, mg/L	
						Primary Effluent	Final Effluent	Primary Effluent	Final Effluent
Jan.	443 (13.4)	0.87	2,323	19.1	9.8	92	3	17.4	4.7
Feb.	642 (14.6)	0.73	2,671	17.3	9.1	69	3	12.7	4.7
Mar.	713 (16.2)	0.53	2,571	11.7	7.8	65	2	9.3	0.4
Apr.	713 (16.2)	0.58	2,333	10.1	8.0	84	3	12.5	0.2
May	726 (16.5)	0.58	2,248	8.3	7.9	87	2	10.8	0.7
June	717 (16.3)	0.50	2,380	10.5	8.1	96	2	11.5	0.1
July	660 (15.0)	0.52	2,584	8.3	8.8	89	2	11.2	0.1
Aug.	660 (15.0)	0.50	2,723	7.5	8.8	85	2	11.7	0.1
Sept.	638 (14.5)	0.58	2,734	15.3	9.1	83	2	11.7	0.3
Oct.	678 (15.4)	0.54	2,410	11.2	8.1	68	2	9.8	0.2
Nov.	634 (14.4)	0.53	2,003	14.8	9.1	72	2	10.6	0.2
Dec.	642 (14.6)	0.49	1,869	14.5	9.1	79	3	10.8	0.2

* Average monthly values are presented.

modification of the activated sludge complete mix or plug flow reactor configuration. Extended aeration plants are operated to maximize endogenous respiration; consequently, solids retention times of 25 to 35 days are not uncommon. Because of their long aeration periods, however, they can suffer from unusual heat losses and low temperatures in cold climate areas. Extended aeration plants can be expected to nitrify fully except at temperatures less than about 8°C because of their characteristic long solids retention time and the resulting accumulation of the slower growing nitrifying organisms. Nitrification in activated sludge systems is reported to occur at temperatures as low as 2°C (36°F) (37).

An oxidation ditch system represents a modification of the activated sludge process in terms of its reactor configuration (see also Section 8.2.5). Wastewater and the reactor biological solids are pumped around an oval pathway which typically is configured as a single channel or concentric multichanneled reactor (Figure 6-7).

Aerators in the form of brush rotors, disc aerators, surface aerators, draft tube aerators, or fine pore diffusers with submersible pumps provide mixing and circulation in the oxidation ditch as well as oxygen transfer. Oxidation ditches typically are designed in the extended aeration mode at hydraulic retention times of greater than 10 hours and solids retention times of 10 to 50 days.

The increased popularity of oxidation ditch systems over the past 15 years can be traced to the operating simplicity and economic attractiveness of this reactor configuration. Oxidation ditches were first installed in the United States in the early 1960s (38). Since that time, the number of oxidation ditches has increased to more than 1,900 in 1991 (39). Although most ditch systems were not designed for nitrification, operation in the extended mode normally ensures that nitrification will occur, provided sufficient alkalinity is present and there is adequate aeration capacity to meet nitrification demands.

6.4.4.1 Design and Performance Information for Extended Aeration Systems

The procedure to follow in designing extended aeration systems for nitrification is equivalent to that outlined for complete mix systems (Section 6.4.2.1). If the mixing conditions within the biological reactor(s) approximate plug flow, refining the procedure to account for plug flow hydraulic conditions is normally not required. This also applies to the design of oxidation ditch nitrification reactors. If the ditch system is designed for carbonaceous oxidation and nitrification only, versus the inclusion of partial or complete denitrification, care must be taken that the aeration component is capable of maintaining a DO of 2.0 mg/L or above throughout the ditch reactor, as recommended in Section 3.3.3. If lower DO conditions are anticipated, selecting the design solids retention time should account for the effect of DO. Equation 6-1 can be modified to include a Monod expression for DO effects as follows:

$$\mu_N = 0.47 \, [e^{0.098(T-15)}] \left[\frac{N}{K_N + N} \right] \left[\frac{DO}{K_o + DO} \right] \quad (6\text{-}22)$$

where:

DO = DO concentration, mg/L
K_o = half-saturation coefficient for oxygen, mg/L

A value for K_o of 1.0 mg/L is acceptable for design purposes. As implied by Equation 6-22, ammonium oxidation will not occur in zones of the oxidation ditch reactor where the DO concentration is very near zero.

Design and performance information for 17 oxidation ditch municipal wastewater treatment plants in the United States has recently been published (39). Although only 12 of the plants were designed for nitrification, the average effluent ammonia-nitrogen ranged from 0.5 to 2.7 mg/L for all 17 plants. The design hydraulic retention time, at the average design flow, was 10 to 34 hours and the design solids retention time, stated for eight of the plants, was 12 to 48 days.
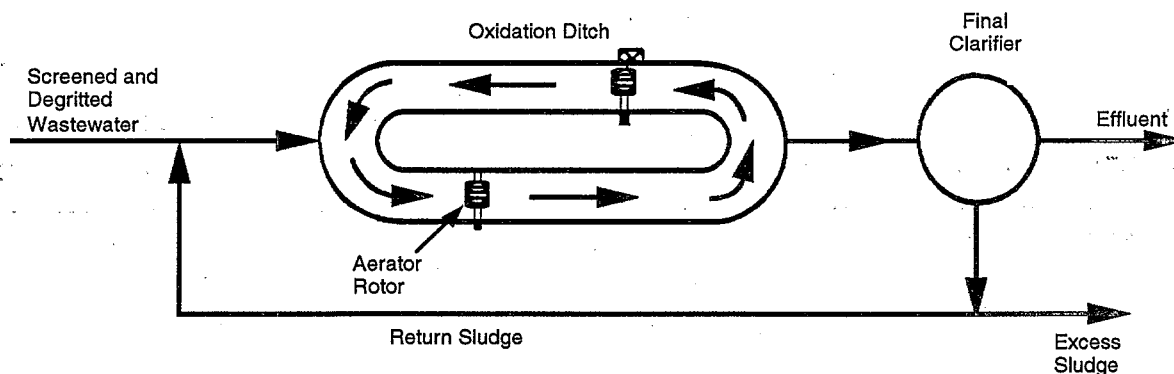

Figure 6-7. Oxidation ditch system.

The Frederick, Maryland, wastewater treatment plant went on line in February 1988. The system consists of primary clarification followed by three parallel oxidation ditch reactors. Aeration is supplied by fine pore· submerged diffusers and each ditch has an intrachannel clarifier. The clarified effluent passes through sand filters before chlorine disinfection and discharge to the Monocacy River. The system is not designed for phosphorus removal beyond that achieved through biomass growth.

The plant is required to achieve a TKN limit of 2.6 mg/L from May through October, the "summer" period. Operating and performance results for the periods of October through December 1990 (winter) and from May through July 1991 are presented along with available design information in Table 6-9. The DO in the oxidation ditch reactors in the summer period typically varied from 1.5 to

2.0 mg/L; maintaining adequate and uniform DO conditions during the warmer months has been difficult on occasion (40).

In winter, the DO varies from 2 to 3 mg/L. The temperature in the oxidation ditch system can be expected to range from 14°C to 25°C (57–77°F) over the year, according to data developed in past years (40). Effluent pH values are typically between 6.8 and 7.2. Although nitrification was evident during the cooler months, the efficiency was significantly less than in the summer, when essentially complete nitrification occurred. Nitrification theory suggests that lower NH_4^+-N levels should be achievable in the oxidation ditch reactors during these lower-temperature months, assuming operation under nonlimiting DO and pH conditions. Incomplete mixing and non-uniform DO and/or pH conditions in the reactor,

Table 6-9. Frederick, Maryland, Wastewater Treatment Plant Design Information and Operating Performance Results (Adapted from Reference 35)

DESIGN

Influent Conditions m³/d (mgd)

Average daily flow	26,500 m³/d (7.0 mgd)
Peak hourly flow	61,300 m³/d (16.2 mgd)

Oxidation Ditch Reactors

Number of oxidation ditches	3
Hydraulic retention time,[a] hr	12.5
Intrachannel clarifier retention time, hr	3.5
Clarifier hydraulic loading rate, m³/m²/d (gpd/sq ft)	25 (600)

Effluent Limits,[b] mg/L	Summer[b]	Winter
BOD$_5$	8.7	26
TSS	26	26
TKN	2.6	(no limit)

PERFORMANCE[c]	Summer (May, June, July 1991)	Winter (Oct., Nov., Dec. 1990)
Influent		
Average daily flow, m³/d (mgd)	27,600 (7.3)	32,900 (8.7)
Primary effluent BOD$_5$, mg/L	116	127
TKN, mg/L	25.9	22.1
NH_4^+-N, mg/L	16.8	14.4
Effluent		
BOD$_5$, mg/L	5.6	14.5
TSS, mg/L	3.5	10.8
TKN, mg/L	2.1	4.7
NH_4^+-N, mg/L	0.53	2.8
NO_x-N, mg/L	20.8	14.8

[a] Calculated at average daily flow.
[b] Summer is defined as May through October.
[c] MLSS varied from 2,000 to 2,500 mg/L, and F/M from 0.12 to 0.15.

sometimes a characteristic of oxidation ditch systems, may account for this variation.

6.4.5 Contact Stabilization Systems

The contact stabilization system can be considered a modification of the activated sludge process in terms of reactor configuration. Return activated sludge is separately aerated in a sludge reaeration or stabilization tank prior to mixing with the influent wastewater (Figure 6-2). Backmixing between the contact tank and the sludge reaeration tank is prevented by providing overflow weirs or pumps between the tanks. The contact tank has a relatively short detention time, 0.5 to 1 hour based on average dry weather flow (ADWF). $CBOD_5$ removal is accomplished in the contact tank primarily by adsorption. $CBOD_5$ removals can be high because the bulk of the organics in municipal wastewater are particulate or colloidal in nature and can be adsorbed to the biological solids for later oxidation in the sludge reaeration tank. The system is not well suited for complete nitrification. Although the overall system solids residence time can be long, the effective wastewater contact time for nitrification is normally insufficient to achieve complete nitrification. Insufficient biological mass is present in the contact tank to nitrify the ammonium completely and since ammonium is not adsorbed on the biological floc, ammonium will bleed through to the effluent. Particularly unfavorable conditions for nitrification will result during peak hydraulic and/or organic loads. Although the contact stabilization reactor configuration is not appropriate for achieving complete nitrification, the system can be considered for partial nitrification, particularly if there are seasonal requirements (41,42).

6.4.6 Step-Feed, Tapered Aeration, and Sludge Reaeration Systems

The step-feed or step aeration system differs from a conventional plant in that influent wastewater is introduced at several points along the aeration tank (Figure 6-2). This distribution of influent flow reduces the initial oxygen demand often experienced in the conventional plant, where mixing approaches plug flow conditions. A variation to the step-feed system involves introducing no feed into the first pass while directing the flow into the remaining downstream passes, creating a sludge reaeration zone in the first pass. This "sludge reaeration plant" is similar to a contact stabilization process except that the contact times are normally longer. Generally, no provision is made to prevent backmixing between the sludge reaeration zone and the downstream aerated zones. The River Oaks plant, discussed in Section 6.4.3.3, represents a system where sludge reaeration occurs in a tank external to the carbonaceous oxidation-nitrification reactors.

Advantages that are claimed for the step-feed activated sludge process include production of a better-settling sludge (43) and the flexibility to vary the solids concen-

tration near the effluent end of the aeration tank while maintaining a constant reactor solids retention time (44,45). The ammonia bleedthrough characterizing contact stabilization plants is avoided in step-feed systems because of the greater contact times employed and because backmixing of the influent typically occurs.

A process modification to conventional or plug flow activated sludge systems that regulates the oxygen supply along the length of the reactor is referred to as tapered aeration. Operational control is better facilitated when tapered aeration is incorporated into conventional systems.

6.4.6.1 Design and Performance Information

The kinetics of a step-feed system are often better described assuming complete mix versus plug flow hydraulics because of the feed pattern and the occurrence of backmixing. Depending on the tank configuration, however, some step-feed plants have plug flow characteristics in each pass. The design approach developed for complete mix systems (Section 6.4.2.1) can usually be employed for step-feed plants as a reasonable approximation. The approach essentially allows determination of the design solids retention time for achieving a required effluent ammonium level. The step-feed process configuration in combination with sludge reaeration has been shown to accommodate a reduction in aeration tank size by allowing an increase in the total mass of solids under aeration (45).

Tapered aeration is commonly employed in the design of conventional or plug flow activated sludge systems. The nitrification design approach developed for these systems (Section 6.4.3.1) is applicable here. Historically, the ability to regulate oxygen supply to the aeration tanks made the tapered aeration configuration attractive when nitrification was not desired. This feature is no longer relevant in the context of designing for nitrification, but tapered aeration does provide a solution to the problem (described in Section 6.4.3.1) of supplying sufficient DO to the head end of plug flow aeration tanks.

As mentioned earlier, there is a danger that if influent is fed too near the effluent discharge point in the step-feed process, insufficient time will be available for organic nitrogen hydrolysis. Under such conditions elevated levels of organic nitrogen could be observed in the effluent. This phenomenon was identified as contributing to the high concentration of effluent organic nitrogen observed during a test study completed at the Flint, Michigan, municipal treatment plant a number of years ago (46).

The Flint test study was conducted in support of a plant upgrade to comply with state regulatory requirements for nitrification. At the time of the test, the existing plant had three aeration tanks, each with four passes, providing a total capacity of 21,225 m³ (750,000 cu ft). With an average design BOD_5 loading of 11,110 kg (24,500 lb)/d to

the aeration tanks at a 75,700 m³/d (20 mgd) flow, the aeration tank load was 523 g/m³/d (32.7 lb/1,000 cu ft/d). Flows to the facility were varied, however, to provide a variation in loading. Three secondary sedimentation tanks were provided with a design overflow rate of 27.6 m³/m²/d (678 gpd/sq ft) at ADWF conditions. The plant was usually operated in a step-feed mode, with one-half the influent directed to the head ends of the second and third passes. During the test, which lasted for a period of 10 months, ferric chloride and polymer were added to the primary treatment stage for phosphorus removal. This addition also had the effect of reducing the organic loading to the aeration tank.

Performance information during eight months of the test are shown in Table 6-10. While nitrate and nitrite are not shown, it was reported that a relatively good balance between ammonium removal and nitrate production was obtained. Nitrite nitrogen was always less than 0.1 mg/L. The appearance of high concentrations of organic nitrogen was attributed to the low rate of hydrolysis of organic nitrogen compounds, likely caused in part by the step-feed process configuration. The 24.1 mg/L of effluent SS also contributed to the elevated effluent levels of organic nitrogen.

Table 6-10. Average Nitrification Performance at Flint, Michigan, for Eight Months (Adapted from Reference 46)

Parameter	Raw Waste-water	Settled Waste-water	Secondary Effluent
BOD₅, mg/L	250	131	13.6
SS, mg/L	300	140	24.1
TKN, mg/L	27.6	23.3	7.8
Organic nitrogen, mg/L	13.3	9.9	6.1
Ammonia nitrogen, mg/L	14.3	13.4	1.7
Phosphorus, mg/L	15.4	2.7	2.3
Temperature, °C	7.2–18.3		

The effect of temperature and solids residence time (θ) are considered in Table 6-11. Effluent quality deteriorated somewhat with colder temperatures, with only 75-percent ammonium removal at 10°C (50°F), at a solids residence time of six days. This drop is not surprising considering nitrification theory (Equation 6-1) which would predict that a minimum solids residence time of approximately seven days would be required to achieve essentially complete nitrification.

6.4.7 High-Purity Activated Sludge System

The use of high-purity oxygen versus air for aeration as a process modification to activated sludge reactors was first evaluated in the 1950s (47). Both covered and un-

Table 6-11. Effect of Temperature and Solids Residence Time on Nitrification Efficiency at Flint, Michigan (Adapted from Reference 46)

Temperature, °C	θ, d	NH_4^+-N Removal, %
18 and greater	4	95
13	4–5	87
10	6	75
7	10–12	50*

* Based on bench scale test results.

covered reactors have since been used but only the former technique has seen wide application at the full-scale level. The covered reactor approach involves the recirculation of reactor off-gases to achieve efficient oxygen utilization. The system uses a three- or four-staged oxygenation basin for contact of gases and mixed liquor (Figure 6-8). High-purity oxygen ($\geq$90 percent purity) enters the first stage and flows concurrently with the wastewater being treated. The gas continues to be used in successive stages, resulting in the buildup of carbon dioxide (released by biological activity) in the gas and in the liquid.

Advantages that are claimed for the high-purity oxygen activated sludge systems, when nitrification is required, include assurance that no DO rate limitations will occur (48) since reactor DO values are typically greater than 4 mg/L (49), and reduced reactor hydraulic retention time requirements due to operation at a higher MLSS concentration (48). Reviews of plant information have indicated that in some cases oxygen activated sludge systems had reduced energy requirements, produced a better settling sludge, and provided a greater tolerance for peak organic loadings than conventional air activated sludge systems (49).

The buildup of carbon dioxide in pure oxygen systems typically causes a pH depression, and nitrification can cause a further pH reduction. While the pH is also depressed by biological carbon dioxide release in conventionally aerated systems (see Section 3.3.4), the pH depression is less significant than it can be in pure oxygen systems because the evolved carbon dioxide is continually stripped from the system by the aeration air. Further information on the effect of the aeration system on the reactor pH is provided in Section 6.4.10.3.

6.4.7.1 Design and Performance Information

The procedure to follow when designing pure oxygen based nitrification systems is equivalent to that for complete mix (Section 6.4.2.1) or conventional (Section 6.4.3.1) systems, depending on the hydraulic design of the plant. Historically, the pure oxygen based systems have been designed as two separate stages, with each
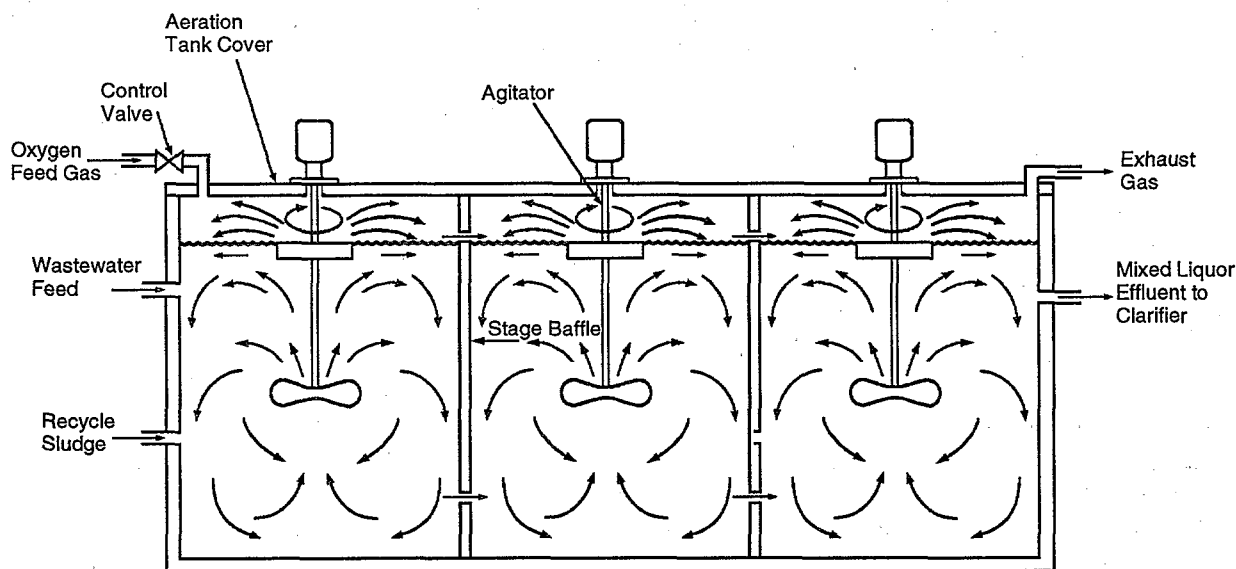
Figure 6-8. Covered high-purity oxygen reactor with three stages and mechanical aerators.

stage consisting of three or more completely mixed reactors operating in series.

If pH conditions in the nitrification reactor(s) are anticipated to be below or near the low end of the optimum pH range for biological systems (this range was suggested as 6.5 to 8.0 in Section 3.3.4), the effect of pH should be considered in selecting the design solids retention time for high-purity oxygen systems. Equation 6-1 can be modified as follows:

$$\mu_N = 0.47[e^{0.098(T-15)}]\left[\frac{N}{K_N + N}\right][1 - 0.833\,(7.26 - pH)] \tag{6-23}$$

where:

pH = anticipated pH in nitrification reactor(s)

The pH effect modification was proposed by Downing and Knowles (50) and further cited in the 1975 EPA Design Manual (1). Understand that acclimation or organism population selection will attenuate the effect of pH on the nitrification rate, as has been discussed in Section 3.3.4. The pH correction term in Equation 6-23 can be considered more applicable to unacclimated cultures within the pH range of 6.1 to 7.2, and should be considered conservative when applied to acclimated cultures.

Widespread utilization of oxygen activated sludge systems for municipal and industrial wastewater treatment began in the 1970s. In 1976 the results of a comprehensive study on the effect of pH on pure oxygen nitrification systems were reported (51). The study, completed at the EPA-DC Blue Plains treatment plant, consisted of two carefully controlled pilot investigations involving nitrification of a wastewater with a low C:N in:

- four stage, high-purity oxygen activated sludge systems with and without pH control, and

- four stage, high-purity oxygen and conventional air activated sludge systems controlled at the same pH level.

Detailed information concerning operating conditions and performance results is provided in Reference 51. The oxygen activated sludge system with pH control produced a mean effluent ammonium-nitrogen level of 0.2 mg/L. The effluent value for pure oxygen systems without pH control was normally equal to or less than 1 mg/L, and was more variable than observed for the pH controlled system. In the controlled system, the pH was maintained at approximately 7.0 by lime addition to the first stage, while in the uncontrolled system pH conditions of 6.0 or below were observed in the last stage. The fact that nitrification was only minimally affected by the low pH conditions is likely a result of two factors: first, the solids residence time in the uncontrolled system was well above the minimum value required (i.e., θ_c^m); second, it is likely that acclimation and/or population selection resulted in the nitrifiers operating at rates approaching those observed under neutral pH conditions. The concentration of organics and nitrogen species in the effluents were virtually identical during operation of the oxygen and air activated sludge systems at the same pH level. The lime requirements to maintain a pH of 7.0 in the last stage of each four-stage system were 2 to 3 times greater in the high-purity oxygen system.

The Town of Amherst Wastewater Treatment Facility in Amherst, New York, discharges to Tonawanda Creek. The plant was designed in the late 1970s for an average daily wastewater flow of 1,050 L/s (24 mgd), anticipated in the year 1990. The plant began operation in 1980. Carbonaceous oxidation and nitrification, following primary clarifi-

155

cation and flow equalization, are accomplished in two separate reactor-clarifier stages using high-purity oxygen activated sludge. Following the Stage 2 clarifiers, the effluent undergoes sand filtration and seasonal chlorination (i.e., from May 15 through October 15) prior to discharge. Phosphorus removal is accomplished by ferrous sulphate addition to the Stage 1 reactors. The nitrifying sludge inventory is maintained at an appropriate level in the Stage 2 reactors by routing a portion of the equalized, primary treated wastewater directly to the reactors. The wastewater flow bypassed to Stage 2 is typically 198 L/s (4.5 mgd) but ranges from 0 to 308 L/s (0 to 7.0 mgd).

Design information pertaining to the carbonaceous oxidation-nitrification system of the Amherst plant is summarized in Table 6-12. Operating and performance information over the 12-month period from October 1990 to September 1991 is presented in Table 6-13. The DO in the Stage 1 and Stage 2 reactors during the period was quite high, typically exceeding 10 mg/L. The pH decrease across the plant, based on raw wastewater and final effluent determinations, was typically 0.4 units. The effluent pH ranged from 6.2 to 7.3. The final effluent alkalinity ranged from 90 to 174 mg/L (as $CaCO_3$). The temperature in the reactors during the coldest months

Table 6-12. Town of Amherst, New York, Wastewater Treatment Plant Carbonaceous Oxidation-Nitrification System Design Information Following Primary Clarification (Reference 52)

Design Flow	90,840 m³/d (24 mgd)
Carbonaceous Oxidation Reactors (Stage 1)[a]	
Number of parallel trains	2
Number of reactors in each train	3
MLTSS, mg/L	6,000
MLVSS, mg/L	3,950
Hydraulic retention time, hr	1.05
F/M, kg BOD_5/kg MLVSS/d	1.17
Stage 1 Clarifiers[b]	
Number	4
Hydraulic loading rate, m³/m²/d (gpd/sq ft)	18.3 (450)
Solids loading rate, kg/m²/d (lb/sq ft/d)	112 (23)
Nitrification Reactors (Stage 2)[a]	
Number of parallel trains	4
Number of reactors in each train	3
MLTSS, mg/L	6,000
MLVSS, mg/L	4,450
Hydraulic retention time, hr	2.10
θ, d	23.8
F/M, kg BOD_5/kg MLVSS/d	0.09
Stage 2 Clarifiers[b]	
Number	4
Hydraulic loading rate, m³/m²/d (gpd/sq ft)	15.9 (390)
Solids loading rate, kg/m²/d (lb/sq ft/d)	98 (20)

[a] Hydraulic retention time calculated at design flow.
[b] Hydraulic and solids loading rates to clarifiers based on design flow.

(i.e., January, February, and March) was estimated between 10°C (50°F) and 11°C (52°F). The performance results (Table 6-13) indicate a slight deterioration in ammonium oxidation during February and March when the temperature in the reactors was near 10°C and the Stage 2 solids residence time was between 10 and 13 days. The effluent pH typically was 6.7 to 6.8 during these months. To achieve an effluent NH_4^+-N of 1.0 mg/L or less at 10°C (50°F) in a complete mix reactor, nitrification theory (Equation 6-23) would predict that a minimum solids residence time of 12 days would be required, if K_N was 1.0 mg/L, the pH was 6.7, and operation was under nonlimiting DO conditions. For the staged oxygen reactor (i.e., complete mix zones in series) an SRT of less than 12 days would be predicted to be adequate.

6.4.8 Sequencing Batch Reactor Systems

Sequencing batch reactor (SBR) systems represent a modification of the activated sludge process. SBRs include a generic system of variable volume activated sludge units in which aeration, sedimentation, and decant are combined in a single reactor. Consequently, there are no dedicated secondary clarifiers or associated return sludge facilities.

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 IFID reactor treats the aqueous wastewater feed stream through a succession of operating steps, namely, fill, react, settle, draw, and idle (Figure 6-9). The liquid volume inside the SBR increases from a set minimum volume to a predetermined maximum volume during the fill period. Mixing and/or aeration may be provided during this first step to promote biomass growth, organics oxidation, and nitrification-denitrification. During react, flow to the tank is discontinued and aeration and/or mixing are provided, while sufficient time is allowed for the microbial reactions to take place. During settle, quiescent conditions are initiated and the biomass is allowed to flocculate and settle prior to removal of the treated clarified supernatant from the tank to the minimum volume level. During the idle period, which is a normal component in multireactor installations, biomass is retained in the reactor but no waste is treated. During this period, excess biomass may be removed from the tank to maintain a desired reactor solids residence time.

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,

Table 6-13. Town of Amherst, New York, Wastewater Treatment Plant Carbonaceous Oxidation-Nitrification System Operating Conditions and Performance Results, October 1990 to September 1991 (Reference 52)*

Month	Wastewater Flow L/S	Wastewater Flow mgd	MLTSS, mg/L Stage 1	MLTSS, mg/L Stage 2	HRT, hr Stage 1	HRT, hr Stage 2	θ, d Stage 1	θ, d Stage 2	BOD₅, mg/L Influent	BOD₅, mg/L Stage 1 Effluent	BOD₅, mg/L Final Effluent	TKN, mg/L Influent	TKN, mg/L Stage 1 Effluent	TKN, mg/L Final Effluent
Oct.	902	20.5	7,475	5,183	1.2	2.4	7.9	18.2	110	22	1	22.1	8.9	0.6
Nov.	871	19.8	8,298	5,063	1.3	2.5	6.3	16.3	119	11	2	22.6	10.5	0.8
Dec.	1,183	26.9	7,358	4,721	0.9	1.9	5.2	12.0	94	14	2	16.9	9.1	0.7
Jan.	1,016	23.1	8,330	5,174	1.1	2.2	7.3	13.7	101	14	2	18.4	10.9	0.9
Feb.	1,007	22.9	7,492	5,707	1.1	2.2	7.1	13.0	102	13	4	18.4	11.5	1.9
Mar.	1,197	27.2	6,487	5,328	0.9	1.9	4.8	9.9	82	13	3	16.1	8.9	2.8
Apr.	1,042	23.7	7,826	4,786	1.1	2.1	7.1	16.0	94	7	2	17.4	8.5	1.1
May	849	19.3	9,357	6,279	1.3	2.6	9.1	10.8	124	25	2	24.3	16.6	0.7
June	765	17.4	8,057	6,695	1.4	2.9	9.6	19.4	122	13	1	24.6	14.5	0.7
July	770	17.5	8,442	5,881	1.4	2.9	12.4	19.2	122	12	2	24.2	14.6	0.8
Aug.	752	17.1	7,361	4,454	1.5	2.9	11.0	10.0	125	36	1	24.9	15.5	0.6
Sep.	756	17.2	5,236	3,202	1.5	2.9	7.7	9.1	147	14	1	28.3	9.5	0.7

* Average monthly values are presented. Flow rate given includes plant-process-water return flow. Hydraulic retention time based on stated wastewater flow. Influent represents the raw wastewater prior to grit removal. Final effluent represents the effluent after chlorination (when required).
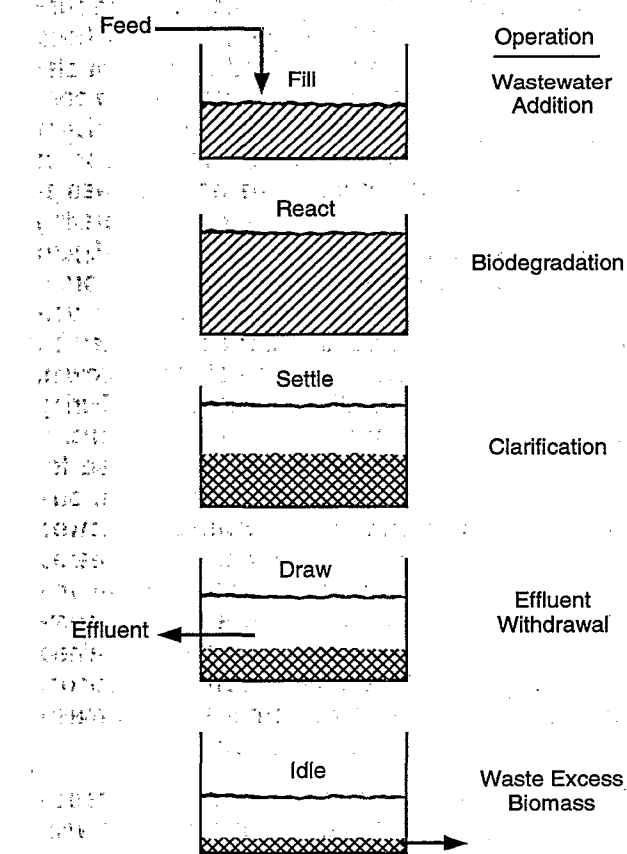

Figure 6-9. Single-tank SBR system operating steps.

using rectangular reactors with length-to-width ratios of at least 2 to 1 and providing baffling.

Major advantages that are often cited for the SBR technology include the ability to tolerate peak flows and shock loads of BOD₅, no need for separate clarification and return sludge pumping systems, and controlled effluent discharge (53). The first modern, full-scale plant for SBR treatment of municipal wastewater in the United States was the Culver, Indiana, wastewater treatment facility. Retrofitted for the SBR process, operation was initiated in May 1980 (54). Since that time, SBR technology has become widespread in the United States, with more than 150 plants in design or operation (55). SBRs can be modified to provide carbonaceous oxidation, nitrification, and biological nutrient removal. Approximately 25 percent of all SBR systems currently operating were designed to achieve nutrient removal (53).

The SBR technology is particularly attractive for treating smaller wastewater flows. The majority of the plants operating to date were designed at wastewater flow rates of less than 22 L/s (0.5 mgd) (55). The cost-effectiveness of SBRs may limit their utilization to flows less than 440 L/s (10 mgd) (54). 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.

6.4.8.1 Design and Performance Information

Each supplier of SBR system equipment has their own approach to design (see Section 8.2.6.1). Some SBR systems are custom designed and the uniqueness of each of these systems reflects the preferences of the design engineer. Designs include the use of different tank configurations, different system hydraulics and a variety of options for aeration, mixing, effluent discharge, and sludge wasting. Systems are normally configured to vary their operation automatically in response to changes in influent flow rate, or to allow the operator to initiate changes to the total cycle time or individual step times, or to make changes during each step (e.g., change length of time for aeration or mixing during fill step). 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 6-14. Nitrification takes place during the react phase and during the portions of the fill period when aeration is practiced.

In order to design SBRs for nitrification, an adaptation of the approach used in the design of complete mix systems is normally acceptable. The specific calculation procedure will be dictated by the characteristics of the selected SBR system. The most important calculation steps are to determine the minimum required aerobic solids residence time (using Equation 6-1 or the modified versions discussed earlier, which account for the effects of limiting DO and pH conditions), and to determine the minimum volume requirements that will assure adequate time for settling and decanting. Other critical parameters for the design of the SBR system can be determined from information presented in Section 8.2.6.2 and elsewhere (55).

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/d). Consequently, partial or complete nitrification is nearly always observed (53,55). In a recent evaluation of 19 SBR treatment plants (53) (all originally designed for nitrification), influent and effluent ammonia-nitrogen data were reported for eight of the plants (Table 6-15). 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.

6.4.9 Powdered Activated Carbon Activated Sludge Systems

The powdered activated carbon (PAC) activated sludge system is a process modification of the activated sludge process. The addition of PAC to plug flow and complete mix suspended growth reactors is a more common process modification for industrial wastewater treatment, and has been applied to municipal systems in some instances. PAC is added to the aeration tank, where it is mixed with the biological solids (Figure 6-10). The mixed liquor solids are settled and separated from the treated effluent in a gravity clarifier. Polyelectrolyte will normally be added prior to the clarification step to enhance solids-

Table 6-14. Typical Cycle for a Single Tank in a Dual Tank SBR System Designed for Nitrification (Adapted from Reference 55)

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 hours (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 hour (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 hour (variable)	Decant—remove clarified effluent from reactor; 15 to 25 percent 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)	Multi-tank 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

Note: A typical total cycle time is 4 to 6 hours.

Table 6-15. Nitrification Performance Information for SBR Operating Plants (Adapted from Reference 53)*

Plant Location	Period of Evaluation	Wastewater Flow m³/d	Wastewater Flow mgd	Percent of Design Flow	BOD₅, mg/L Influent	BOD₅, mg/L Effluent	Ammonia-N, mg/L Influent	Ammonia-N, mg/L 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	2,176	0.575	72	195	4	15.8	1.2
Marlette, MI	07/90–06/91	1,578	0.417	60	103	4	10.1	0.5
Mifflinburg, PA	10/88–03/91	2,763	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	2,116	0.559	56	160	7	12.9	0.6

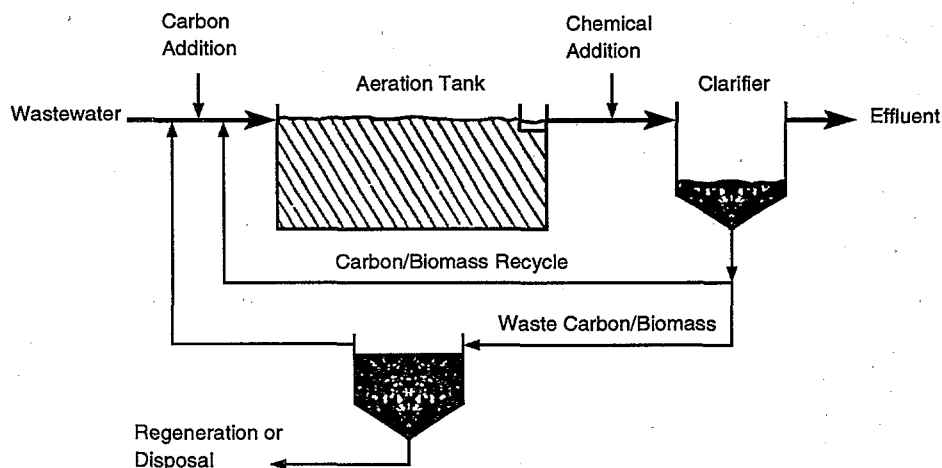* Average monthly values based on all data available.

Figure 6-10. Powdered activated carbon activated sludge system.

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 SS. The clarifier underflow solids are 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.

Demonstrated advantages of PAC addition to suspended growth reactors include improved solids settling and dewatering characteristics; the ability of PAC to adsorb biorefractory materials and inhibitory compounds, improving effluent quality and reducing the impact of organic shock loads; reduction in odor, foaming, and sludge bulk-

ing; and improved color and CBOD₅ removal (49). 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 (2,400–4,000 lb)/d, wet air oxidation/regeneration (WAR) is claimed to represent an economical approach to carbon recovery and waste biomass destruction (56). However, an ash separation step is needed in this case, affecting the economics of carbon regeneration and recovery (57). 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 (58,59), the economics of complete PAC/WAR systems relative to other activated sludge nitrification systems are unclear (57).

In the United States, PAC activated sludge 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 (60). A summary of selected municipal PAC facilities is presented in Table 6-16.

6.4.9.1 Design and Performance Information

The procedure to follow in designing PAC activated sludge systems for nitrification involves a modification to those for complete mix (Section 6.4.1) or conventional plug flow systems (Section 6.4.3.1) in order to account for the effects of the addition of PAC. According to the major supplier of the technology (60), most PAC systems are designed at MLTSS 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 (60). 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. The PAC concentration to be added can be calculated from:

$$PAC_o = \frac{(PAC_R)t}{\theta_c^d} + PAC_1 \qquad (6-24)$$

where:

PAC_o = influent PAC concentration, mg/L
PAC_R = mixed-liquor PAC concentration in the reactor, mg/L
PAC_1 = effluent PAC concentration, mg/L
t = hydraulic retention time, d

The value of PAC_1 in Equation 6-24 can be estimated by assuming that the carbon fraction in the effluent TSS is the same as the fraction of PAC in the MLSS.

As an example, if complete mix hydraulics were employed for the bioreactor of the PAC activated sludge system, the design procedure would follow Steps 1 through 8 as detailed in Section 6.4.2.2. In order to complete Steps 9 and 10, X needs to be selected recognizing that the total MLSS will now include PAC. Once X and PAC_R are specified and t is determined from Equation 6-15, the required influent PAC concentration can be calculated from Equation 6-24.

PAC activated sludge 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, as noted in Section 3.3.5 and

Table 6-16. Summary of Municipal PAC/WAR Facilities Reviewed (57)

Facility	Current/Design Flow, m³/s	PAC/WAR[a] Status	Reason for PAC[b]	BOD₅, mg/L	TSS, mg/L	NH₄⁺-N, mg/L
				Permit Limits		
Vernon, 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. Olmsted,[c] OH	0.26/.31	AS	N,S	30	30	2.3–6.9
Sauget, IL	0.70/1.2	AS	T	20	25	—
El Paso, TX	0.20/.44	MA	N,O	SD[d]	SD	SD

[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.

in Section 6.3.1. Researchers have provided evidence that the addition of PAC to nitrifying activated sludge systems receiving industrial wastewaters improved nitrification rates (61,62). More recently, 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 (63). 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.

6.4.10 Other Design Considerations for Suspended Growth Nitrification Systems

6.4.10.1 Selection of Peaking Factors, Safety Factors, and Process Design Factors

The selection of peaking factors and safety factors for process design should not be confused. Peaking factors are used to reflect assumed realities under the controlling conditions of the design. Safety factors are used to reflect uncertainty in performance under these realities. Whether or not the two are conceptually multiple to establish an overall process design factor depends on the application; judicious application of both peaking and safety factors can avoid a clearly inappropriate under- or over-designed condition.

The development of the design example in Chapter 2 introduces the reader to concepts associated with the development of the peaking factors for process design. They first reflect the compliance interval of the plant's effluent objectives (Section 2.9.3.1). These factors then consider the variability of the influent wastewater characteristics (Section 2.9.3.2) and their impact on the integrated works of the treatment plant (Section 2.9.3.3) for the processes under investigation (Section 2.9.3.4). The impact of the wastewater characteristics and their implied peaking factors are evaluated under the planned modes of operation through the preparation and use of mass balances (Sections 2.9.3.5 and 2.9.3.6, respectively). The developmental discussion and concluding table for the design example (Table 2-18) show that the elected processing peaks vary as a function of the unit process and processing considerations, with the layered impact of the processing recycles.

The designer can influence the process design peaking factors for the raw wastewater and processing recycles through equalization and diversion, or split-treatment strategies, and the elected processes and operating strategies for the treatment works. Longer hydraulic and solids residence times, and continuous, as opposed to discontinuous, processing, serve to mitigate pollutant mass peaks through the treatment system.

Flow peak mitigation is one of the most important—if not the most important—concerns in suspended growth systems because of solids washout concerns. Here, the designer should consider the applicability of automatic high flow diversions around the reactor and/or the entire biological treatment system, as well as flow equalization with the equalization tank dedicated to receiving both the first and the end-of-storm sewer flushing events. Other peaking factor considerations, beyond these general considerations, are best considered as a function of the reactor design.

Safety factors in process design are ultimately expressions of design confidence. They are used when there is uncertainty. Higher safety factors may be used where the technology is less demonstrated or more unproven. One example of a commonly used safety factor in design is multiplication by two of the reciprocal of the controlling design Nitrosomonas specific growth rate (μ_N). This solids residence time is expected to accommodate unknown variations in temperature, DO, residual ammonium concentration, operating variations, and pH (if determined to be applicable). Conceptually, the sound design approach would be then to apply this solids residence time under the controlling design conditions that are established by correct application of the peaking factors.

6.4.10.2 Incorporation of Phosphorus Removal

Chapter 2 provides the stoichiometry of the phosphorus removal reactions (Table 2-3), and the considerations associated with the incorporation of phosphorus removal with a suspended growth nitrification system are largely developed in Section 2.7.2 and demonstrated in Design Example No. 1 (Section 6.4.2.2). A summation of the salient points as they may influence suspended growth nitrification systems follows:

- Solids Production:
  - All designs must anticipate the production of additional waste solids due to the phosphorus removal step and select a design MLSS concentration with this in mind.
  - Stringent levels of phosphorus removal may not be obtainable with secondary equivalency levels of effluent SS (e.g., maximum 30-day average of 30 mg/L) because of the increasing phosphorus concentrations in the sludge mass and effluent SS.

- Metal Salt Addition Strategies:
  - Must anticipate the attendant alkalinity depletions if an acid carrier is used with the metal salts.
  - May be especially attractive in multipoint addition strategies for improved solids-liquid separation (with or without polymers), and with upstream primary

clarification as a means of minimizing reactor requirements for cold temperature nitrification, as well as lower overall dosages.

- Are strongly suggested as backup (and planned utilization) for strategies that anticipate biologically enhanced phosphorus removal strategies to ensure compliance with the plant's effluent objectives.

- Biologically Enhanced Phosphorus Removal Strategies:

  - Must pay particular attention to over-aeration (the best performance is not expected with nitrifying systems), mainstream solids separation, and the solids processing train to minimize the likelihood of phosphorus resolubilization and return.

  - May be best applied with the direct and immediate application of lime to the waste sludge stream (high pH phosphorus precipitation) to ensure the permanent removal of the phosphorus contained in the waste solids beyond normal synthesis or metal precipitation expectations.

6.4.10.3 Effect of Aeration System on Process Design

The practical realities of the elected aeration system have a strong influence on process design in that they can influence the DO and the pH in the system, as well as the character of the effluent SS. The first two environmental variables can, and do, affect the overall required solids residence time for nitrification. The character of the effluent SS may be altered by the mixing energy used in the oxygen dissolution system.

As explained in Section 2.5.1, the pH of any system reflects its bicarbonate-carbonate equilibrium with the carbon dioxide in the liquid and the atmosphere. Besides the alkalinity demand of the nitrification reaction, variables include the water surface exchange with the atmosphere and the carbon dioxide generated by biological stabilization of the carbonaceous material. For a biological system with a nonlimiting alkalinity, lower pH values can be expected to be promoted by, in order of increasing significance, high MLSS concentrations, high loadings of biodegradable carbon, low physical surface exchange, and whether the reactor surface is open or closed to the atmosphere. Open atmosphere operation and isolation of the nitrification reaction are certain to yield the most favorable pH conditions because of the relative absence of any appreciable carbon dioxide release by the biomass.

The closed pure oxygen activated sludge systems, designed for combined carbonaceous and nitrogenous material oxidation, can, and do, experience low reactor pH conditions (7 or less) and in the absence of pH control provisions, are most safely designed for a longer solids residence time than air systems since rate increases from their elevated DO level may not be completely compensating for the rate decreases resulting from the lower pH.

Thereafter, of probable minor importance, higher pH conditions would be expected from a coarse rather than a fine bubble diffused air supply. With mechanical aeration systems, the resultant pH will vary as a function of the system; total reactor coverage as opposed to the single site application (as classically encountered in ditch concepts) would be expected to result in the most favorable pH conditions.

DO levels are probably the most important influence on the overall solids residence time of the system. Systems that anticipate DO levels below the nonlimiting condition must adjust the design solids residence time upward to reflect the very real slowdown in the nitrification replication rate. This consideration becomes particularly important in systems that try to achieve some denitrification in the nitrification reactor, and leads to the often observed compromise of elevated ammonium concentrations with reasonable control of the effluent oxidized nitrogen levels. The compromise undoubtedly accelerates as design conditions are approached. The significance of this consideration varies with the design intent of the reactor and the dissolution system. Manipulation strategies largely vary with how the air or oxygen is introduced, the reactor configuration and, of course, the operating concentration and age of the MLSS.

Coarse or fine-pore diffusion strategies yield the greatest flexibility and assurance of a nonlimiting nitrification solids residence time. Brush or single-point aeration devices in ditches are the opposite extreme and can yield nitrification limiting DO levels for a significant portion of the reactor volume. Other mechanical aeration systems lie between these two extremes and may also result in nitrification limiting DO levels in the early stages of a plug flow design, or find it difficult to achieve elevated DO levels throughout the reactor's content in a complete mix design. Investigations at 23 activated sludge plants in West Germany clearly show that surface aeration systems did not nitrify as well as diffused aeration plants at similar loadings and temperatures and comparable oxygen concentrations (as measured at the surface of the aeration basin) because of the lack of uniformity of elevated DO levels in the aerator (64).

The type of aeration system may also influence the effluent SS through the resulting localized or systemwide turbulence. Higher levels of effluent SS may be encountered with increasing mixing intensity as represented by the root-mean-square velocity gradient, G. In general, G will increase with a declining oxygen transfer efficiency and depend on the nature of the oxygen dissolution system. Fine-pore diffused aeration systems will have lower G values than coarse bubble systems, and mechanical aeration will most often have G values similar to those encountered in coarse systems, with increasingly intense localized areas of mixing based on the number of aeration points. Lower level MLSS concentrations provide a

means of avoiding high oxygen uptake rates per unit volume of reactor (and high G designs). Tapered aeration, including serial staged application of mechanical aeration systems, and external or internal flocculation clarifiers provide a means of floc reformation and recovery when high Gs are unavoidable.

Recent investigations (65), which involved 105 measurements at nine mechanical aeration activated sludge plants and 150 measurements at 15 diffused aeration activated sludge plants, reveal average supernatant sample SS levels of 17 ± 9 and 11 ± 8 mg/L, respectively. Effluent SS may or may not match these values, depending on the distance from the mechanical aeration system, and the shearing/flocculation opportunities in the transport system from the aerator to the point of separation at the sedimentation tank. Optimum G values are reported to be on the order of 70 s^{-1}, with little perceptible change up to a mixing intensity of about 100 s^{-1}; deterioration is likely once G exceeds 150 s^{-1}. Procedures to calculate G values for mechanical or pneumatic mixing systems are readily available (66).

6.4.10.4 Aeration Requirements and Design

Aeration requirements are determined by the procedures described in Tables 2-15 and 2-16 of Chapter 2. Spatial and temporal variations in oxygen demand can be determined through use of mathematical models such as those described in Chapter 5. Additionally, and especially with fine pore-diffusion systems, the designer should check for adequate mixing (G values of 50 to 70 s^{-1}) and, if applicable, blower turndown during the minimum monthly needs of the design life of the facility.

6.4.10.5 Solids-Liquid Separation

The considerations associated with the design of the sedimentation system for a suspended growth reactor are described in Section 2.7.1.1. In combined carbonaceous and nitrogenous oxidation systems, minimal sludge blankets, which are obtained by operating at high return rates with low return sludge concentrations, will ensure freedom from settled sludge denitrification and resultant rising and floating sludge. Other enhancements may include a rapid hydrostatic, full-floor sludge collection system as opposed to a single-point (often a center-well) sludge collection system; although, the attractiveness of this feature has yet to be established with a dilute return sludge concentration (minimal sludge blanket) design and operating strategy.

6.4.10.6 Seasonal Nitrification Considerations

Seasonal nitrification considerations initiate with the realization that the plant must fully meet its effluent objectives for the first full month that they apply. Accordingly, conditions to achieve this compliance must be established in the preceding month.

The likely controlling environmental parameter is wastewater temperature and its determining influence on the required nonlimiting aerobic solids residence time of the nitrifiers. Long hydraulic residence time reactors are likely to have somewhat lower reactor temperatures than the influent wastewater temperature. Surface aerator systems are likely to have colder reactor temperatures than diffused aeration systems. To ensure that a sufficient nitrifier population will be available at the start of the compliance month, the population must be established in the previous month. If the average monthly reactor temperature for the month that precedes the initial seasonal compliance month is lower than the wastewater temperature during the permit period, the temperature for the month preceding the permit period can normally be used safely for design.

After realizing that the proper temperature and DO dependent solids residence time must be in effect to achieve the required degree of nitrification for the first compliance month, the designer has many choices on how to achieve this objective. The following paragraphs describe several representative approaches.

Total System Responsiveness. Assuming that the coldest wastewater temperatures occur during the month before the compliance standard is applicable, simply design the reactor to achieve the desired θ_c^d for that month. This approach is the most conservative, with the least amount of risk; it may also be the most expensive.

Reactor/Clarifier Isolation or Split Treatment. This approach does the same as the first except that it is limited to a fraction of the plant. The plant fraction dedicated to full nitrification is determined by the allowable effluent ammonium release and the mix of non- and fully nitrified process streams. Here, the designer might think of constant flow treatment to the nitrifying portion of the plant and allow the remaining portion of the plant to experience exaggerated swings associated with the balance of the flow to the plant. Constant flow strategies to the isolated nitrification system will also allow higher MLSS concentrations in this system, since the operating level does not have to be established in anticipation of the diurnal and/or seasonal flow peaks. Waste solids from the nitrifying portion of the plant could also be delivered to the rest of the secondary treatment system to seed this system and achieve some favorable ammonium oxidation.

Lower the Applied Suspended Solids (Carbon). If the plant has primary sedimentation, the applied BOD_5 and SS to the nitrifying activated sludge system may be reduced by chemically enhanced primary treatment through the use of metal salts and/or organic polyelectrolytes. Such strategies will yield a lower net biomass for the same solids residence time objective, and may be particularly attractive as an operating capability as the plant approaches design conditions.

Increase the Aeration Intensity. Whether or not this strategy has any value depends on the certainty of a non-limiting aerobic nitrifying solids residence time. The 9,900-L/s (225-mgd) Metro Plant in Minnesota reports that this strategy is successfully used to achieve its seasonal performance objectives (67).

Breakpoint Chlorination Polishing. This strategy was often conceptually reported in the early years of nitrogen control technology as the means of achieving seasonal compliance. It is uncertain how many facilities use this approach. Since it is an all-or-nothing concept, it may be best applied with a split treatment approach to avoid excessive chlorine usage. The designer should review the experience reported for breakpoint chlorination in Section 2.5.2.3 before proposing this concept, and anticipate the probability of attendant dechlorination needs if the concept is applied.

All seasonal nitrification concepts bring the accompanying certainty of elevated nitrite nitrogen excursions and the resultant chlorine demands if this is the designed disinfection process. Section 2.8.5 provides additional information about this problem. Table 2-3 summarizes the stoichiometric demand.

6.5 Nitrification in Attached Growth Reactors

Reactors in which growth occurs on or within a solid medium are termed attached growth, supported growth, or fixed film reactors. A number of attached growth reactor configurations—including trickling filters, rotating biological contactors, aerated biologial filters, and both packed- and fluidized-bed systems—can be designed to incorporate nitrification. Descriptive and design and performance information pertaining to each attached growth configuration is presented in Sections 6.5.2 through 6.5.4.

6.5.1 Application of Biofilm Models to Design

The design of attached growth processes is primarily based on empirical results from the operation of pilot and full-scale systems. Examples of typical design criteria include volumetric loading rate (mass of substrate per unit of reactor volume per time), surface loading rate (mass of substrate per unit of media surface area per time) and hydraulic retention time (residence time of feed material in reactor, normally based on empty bed volume, i.e., total volume occupied by the media). The inability to determine the VSS concentration easily in most attached growth reactors precludes the use of VSS specific removal rates (e.g., mg ammonium oxidized per mg VSS per day) commonly employed in the design of suspended growth reactors. Although the attached growth reactor design relationships are generally empirical, in most cases they are consistent with biofilm models developed based on stoichiometry, Fick's law, and Monod kinetics. The kinetic considerations for attached growth systems, and related diffusional resistance aspects, are discussed in Section 3.4.

The equations that describe the typical biofilm model (Section 3.4) are presented elsewhere (68). Solving the equations requires numerical techniques or the use of pseudoanalytical solutions to approximate the numerical solutions. A graphical, loading curve approach has been developed which allows for sizing of an attached growth reactor without solving the biofilm model (68). The approach involves the development of families of curves representing flux (i.e., mass of substrate per unit of biofilm surface per time) as a function of substrate concentration in the bulk liquid of a completely mixed biofilm reactor or reactor segment. The flux and substrate concentration are expressed as normalized values using appropriate parameters for normalization. The value of these parameters will be dictated by biofilm kinetic and hydrodynamic considerations (e.g., diffusion coefficients). The graphical representations, further simplified by methods proposed by others (69), provide straightforward relationships for substrate flux as a function of substrate concentration under steady state conditions. A steady state mass balance on a completely mixed attached growth reactor or reactor segment, yields the following equation:

$$J_a = \frac{Q(S_o - S_1)}{a\,V} \qquad (6\text{-}25)$$

where:

J_a = substrate flux rate per unit biofilm surface area
Q = volumetric flow rate, volume/time
V = reactor segment volume, volume
S_o = feed substrate concentration, mass/volume
S_1 = effluent substrate concentration, mass/volume
a = surface area of biofilm per unit reactor volume, area/volume

Equation 6-25 can be used directly to calculate required reactor volume, for given values of the wastewater substrate concentration and the effluent substrate concentration. The calculation requires definition of the specific surface area (a) of the support medium and determination of J_a from the appropriate graphical representation. The use of this procedure has been illustrated in the design of an aerated biological filter for $CBOD_5$ removal, assuming in one case completely mixed reactor hydraulics and in the other, plug flow hydraulics. The procedure is readily adaptable to the design of nitrification reactors operating at low $CBOD_5$ feed concentration. Information pertaining to another practical but fundamentally based approach to the design of attached growth nitrification reactors is presented elsewhere (70).

If nitrification is to occur in a biofilm reactor, competition from the heterotrophs for oxygen and space in the biofilm must be reduced. Based on the biofilm model and practice, it has been proposed that the CBOD flux (i.e., J_a in Equation 6-25, using CBOD as the substrate) must be

less than approximately 1 g/m²/d (0.2 lb/1,000 sq ft/d) in order to obtain reliable nitrification in a biofilm. This proposal is based on measuring CBOD as ultimate CBOD (71). Other modeling efforts (72), verified through data from the literature, indicate that for nitrification to be significant, oxygen must penetrate more deeply into the biofilm than biodegradable organic material. In practice this implies that at a bulk liquid oxygen concentration of 3 mg/L, the concentration of organic material must be less than 20 mg/L as CBOD (72). Based on field observations of operating trickling filters, some authors have suggested that the soluble BOD_5 must be less than 20 mg/L before the onset of nitrification, without stating the corresponding bulk liquid oxygen concentration (73). Others claim that significant nitrification can be achieved at bulk liquid soluble COD levels of up to approximately 60 mg/L without stating the biodegradable fraction of the COD value (74).

Perhaps the most important implication of the biofilm model on the design of attached growth reactors for nitrification is that below a certain bulk liquid ammonium and DO level, the model predicts that the specific ammonium oxidation rate will be reduced as a result of limiting substrate diffusion effects, regardless of the significance of competition from heterotrophic growth. That is, the specific ammonium oxidation rate is no longer zero-order; it is dependent on the bulk liquid ammonium or DO concentration. As discussed in Chapter 3 (Section 3.3.1), nitrifiers under these conditions are no longer operating at their intrinsic growth rates. Measured ammonium volumetric and surface removal rates in attached growth reactors treating municipal wastewater with a low soluble CBOD tend to confirm this theoretical implication (75–85). In fact the measured or apparent rates in trickling filters, based on influent and effluent data, appear to be more influenced by substrate effects than would be expected theoretically (84). This may be due to patchy or incomplete film growth at low bulk liquid ammonium concentrations. Under such conditions the active surface area of the media is less than calculated.

A recent attempt has been made to relate the solids residence time in a trickling filter to the surface loading rate, based on the use of results from field observations and empirical design relationships developed by various authors (74). The design information derived from this effort is empirical, and the observations should not be considered design criteria. Generally, they are applicable to attached growth reactors with a hydraulic regime closer to plug flow than complete mix:

- The onset of nitrification will likely not occur unless the soluble BOD_5 surface loading rate is less than 9 g/m²/d (1.8 lb/1,000 sq ft/d).

- Nearly complete nitrification (effluent ammonium-nitrogen of approximately 2 mg/L) will typically be encoun-

tered at a soluble BOD_5 loading rate of approximately 2 g/m²/d (0.4 lb/1,000 sq ft/d).

- Low-surface or volumetric ammonium removal rates are normally encountered in attached growth reactors that are designed to achieve low effluent ammonium-nitrogen concentrations (1.5 mg/L or less). This is due to the difficulty in building up and maintaining a significant nitrifier biofilm in the latter stages of a plug flow fixed film reactor, where the low ammonium concentration limits nitrifier growth.

Therefore, while biofilm modeling can provide significant insight into the design and performance of attached growth reactors operating at a low influent soluble $CBOD_5$, other factors, such as the influence of particulate influent organics and the nonhomogeneous biofilm thickness and density characteristic to such attached growth reactors, complicate application of theoretical models. A reasonable conclusion is that the mechanisms of nitrification in attached growth reactors operating at a higher influent $CBOD_5$ are so complex that design approaches are necessarily empirical.

6.5.2 Trickling Filters and Biotowers

The trickling filter is an aerobic attached growth reactor in which a solid surface medium is used to support biofilm growth. Wastewater is normally introduced at the top of this attached growth reactor through a distribution system and flows or trickles down through the media. The distinction between trickling filters or biotowers and packed-bed reactors, and aerated biological filters and fluidized bed reactors is that in the latter reactors, the hydraulic design is such that the media are submerged in the reactor liquid during normal operation. Although trickling filters are often designed to allow flooding or submergence of the reactor media for maintenance purposes, operation in this fashion is not routine.

Information pertaining to the application of trickling filters and biotowers to municipal wastewater treatment in general can be found in other publications (49,86). The following discussions are more specific to use of the technology for nitrification of municipal wastewaters.

Trickling filter or biotower media traditionally consist of rocks, slag, or synthetic materials. Rock and slag trickling filters generally have 1.2 to 3.0 m (4 to 10 sq ft) of media depth. Plastic media trickling filters are normally constructed at depths of 4.6 to 7.6 m (15 to 25 ft) because of the lighter weight and better ventilation capabilities of the packing (87). Recent advances in the development of plastic and other media with different structural configurations have made this technology more efficient and cost effective. Structural configurations include assemblies of plastic corrugated sheets, random plastic ring structures (e.g., pall rings), horizontal wood-slatted structures, and polyethylene strip media.

Because of their advantages, synthetic media are typically selected for new plant design instead of rock or slag trickling filters. Examples of synthetic media types that are appropriate for nitrification applications are listed in Table 6-17. Corrugated plastic sheet media fall into two broad classifications, vertical and cross-flow configurations (88). As shown in Figure 6-11, vertical media have corrugations that direct flow primarily downward. Cross-flow media consist of ridged corrugated sheets in which the ridges on adjacent sheets are normally set at 45- or 60-degree angles to each other, and are touching where they cross (Figure 6-11). The configuration provides the opportunity for splitting and combining liquid flow. Studies have shown that, when compared to vertical flow media, the cross-flow configuration results in a longer hydraulic retention or liquid contact time (89) and appears to pro-

Table 6-17. Comparative Physical Properties of Example Synthetic Trickling Filter Media Suitable for Nitrification of Municipal Wastewaters

| Media Type | Characteristics | Unit Weight | | Specific Surface Area | | Void Space, % | U.S. Manufacturer |
		kg/m^3	lb/cu ft	m^2/m^3	sq ft/ cu ft		
X-FLO 30	Plastic sheets, cross-flow, corrugated	24–45	1.5–2.8	98	30	>95	American Surfpac Downington, PA
X-FLO 42	Plastic sheets, cross-flow, corrugated	24–45	1.5–2.8	138	42	>95	American Surfpac Downington, PA
BIOdek 19050	Plastic sheets, cross-flow, corrugated, 60-degree angle	24–45	1.5–2.8	138	42	95	Munters Ft. Myers, FL
BIOdek 10250	Plastic sheets, cross-flow, corrugated, 60-degree angle	24–45	1.5–2.8	223	68	95	Munters Ft. Myers, FL
Accupac CF-1900	Plastic sheets, cross-flow, corrugated, 60-degree angle	27–63	1.6–3.9	157	48	95	Brentwood Industries Reading, PA
Accupac CF-1200	Plastic sheets, cross-flow, corrugated, 60-degree angle	36–63	2.2–3.9	226	69	95	Brentwood Industries Reading, PA
Tripac	Random pack media, polypropylene, 1-inch diameter balls	99	6.2	279	85	90	Jaeger Products Houston, TX
Tripac	Random pack media, polypropylene, 2-inch diameter balls	67	4.2	157	48	93	Jaeger Products Houston, TX
Tripac	Random pack media, polypropylene, 3.5-inch diameter balls	53	3.3	125	38	95	Jaeger Products Houston, TX
BIO-PAC	Random pack media, 2-inch diameter balls	53	3.3	102	31	94	NSW Corp. Roanoke, VA
BIO-PAC	Random pack media, 1.5-inch diameter balls	61	3.8	144	44	93	NSW Corp. Roanoke, VA
BIO-PAC	Random pack media, 1-inch diameter balls	72	4.5	180	55	92	NSW Corp. Roanoke, VA
BIO-PAC	Random pack media, 0.62-inch diameter balls	112	7.0	331	101	88	NSW Corp. Roanoke, VA
Sessil 100	Flexible hanging plastic sheets	N/A	N/A	98	30	N/A	NSW Corp. Roanoke, VA
Sessil 130	Flexible hanging plastic sheets	N/A	N/A	138	42	N/A	NSW Corp. Roanoke, VA
Sessil 200	Flexible hanging plastic sheets	N/A	N/A	223	68	N/A	NSW Corp. Roanoke, VA

Vertical Media   Cross-Flow Media
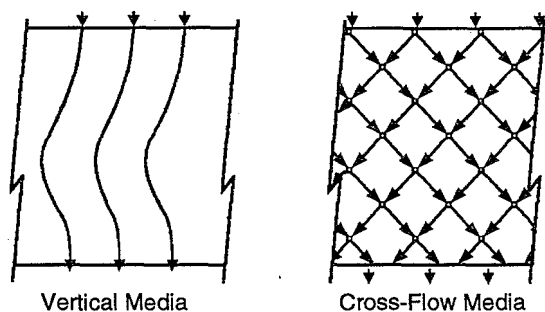
Figure 6-11. Hydraulic flow patterns in vertical and cross-flow trickling filter media (adapted from Reference 88).

vide a higher oxygen transfer efficiency (88). These features allow a substantially higher BOD_5 surface loading rate than can be imposed on vertical media for equivalent effluent quality (88).

Rock or slag media trickling filters can be designed to achieve nitrification by operating them at low organic loading rates, provided that proper ventilation is achieved. A trickling filter or biotower based on the use of horizontal wood slats is currently popular in one commercial embodiment of a combined suspended and attached growth system. Information pertaining to this combined growth reactor system is presented in Section 6.6.2.3. The use of polyethylene strip media in the United States as the packing for a trickling filter or biotower is a relatively new development (90). One example is that currently manufactured by NSW Corporation of Roanoke, Virginia, and referred to as Sessil™ media. The Sessil strip media are 2.9 cm (1.14in) wide and are hung from a wood support structure located at the top of the biotower. The strip media are typically 6 m (20 ft) in length. Operating results from the application of this media to nitrification of municipal wastewaters have yet to be reported.

The design and performance information presented in Sections 6.5.2.1 and 6.5.2.2 is specific to conventional trickling filters employing rock, slag, or plastic media for the treatment of municipal wastewaters at high and low $CBOD_5$, respectively.

6.5.2.1 Nitrification Design and Performance Information at High Carbonaceous Feed Concentrations

The number of operating trickling filters designed to achieve nitrification of municipal wastewater containing a high $CBOD_5$ concentration, typical of primary treated wastewater, is limited. In an assessment completed for EPA in 1991, 10 plants were identified that simultaneously achieved $CBOD_5$ removal and nitrification in a single trickling filter unit, commonly referred to as combined or single-stage carbonaceous oxidation-nitrification (87). Six of the 10 plants used the trickling filter/solids contact

(TF/SC) flowsheet. Note that information pertaining to the TF/SC flowsheet will be presented later in this section.

A degree of ammonium oxidation has been achieved for many years in low or standard rate rock media trickling filters (1). Results from numerous full-scale and pilot plant studies (Figure 6-12) show that for these filters to attain nearly complete nitrification (90-percent ammonium removal), the organic volumetric loading rate must be limited to approximately 80 g $BOD_5/m^3/d$ (5 lb/1,000 cu ft/d). The reduction in nitrification at higher loadings was attributed to domination by heterotrophic bacteria in the microbial film. It was also noted that there were very little data to judge the effect of temperatures on performance below 20°C, but that lower temperatures would require a reduction in the volumetric loading rate. The results indicate some improvement if recirculation around the rock filters is practiced (Figure 6-12). The results from blast furnace slag media trickling filters indicate that comparable volumetric loadings to those of rock filters are necessary to achieve nearly complete nitrification based on results where recirculation is not practiced. Imposition of a recirculation ratio of 1:1 to the slag media filters allowed nearly complete nitrification to be achieved at loading rates over 1.5 times greater than the rock media. The results presented in Figure 6-12 should be considered approximate. In addition to operating conditions differing from site to site, it is not known whether the BOD_5 results reflect $CBOD_5$ or include the effect of ammonium oxidation in the BOD test.

Historically, the rock media used in trickling filters has typically had a nominal size of 2.5 to 10 cm (1 to 4 in) with a corresponding specific surface area of less than 65.6 m^2/m^3 (20 sq ft/cu ft) (49). The higher specific surface areas of plastic media (Table 6-17) have resulted in the ability to accomplish nitrification at higher volumetric loadings, relative to rock media filters. Another factor favoring greater capacity of the plastic media filters is oxygen supply; rock filters often have poor ventilation, particularly when water and air temperatures are close or the same.

In studies completed at Stockton, California (91), good agreement was found between the nitrification performance of plastic media trickling filters and previous results with rock media when the performance was expressed as a function of the BOD_5 surface loading rate (Figure 6-13). The Stockton results indicate that 90-percent ammonium removal was achieved at a BOD_5 surface loading rate (the reference is not specific in differentiating between BOD_5 and $CBOD_5$) of approximately 2.5 g/m^2/d (0.51 lb/1,000 sq ft/d) for both rock and plastic media trickling filters (Figure 6-12). The corresponding BOD_5 volumetric loading rates are 120 g/m^3/d (7.5 lb/1,000 cu ft/d) for the rock media and 220 g/m^3/d (13.7 lb/1,000 cu ft/d) for the plastic media.
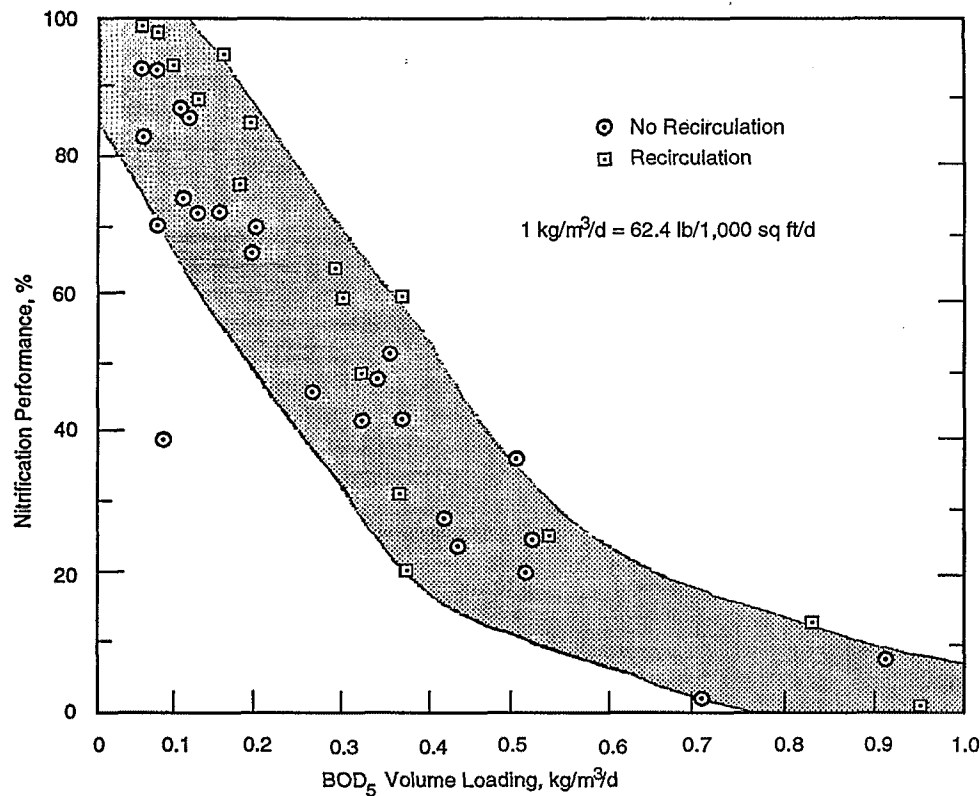
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Figure 6-12. Effect of BOD5 volumetric loading on nitrification performance of rock trickling filters (adapted from Reference 1).
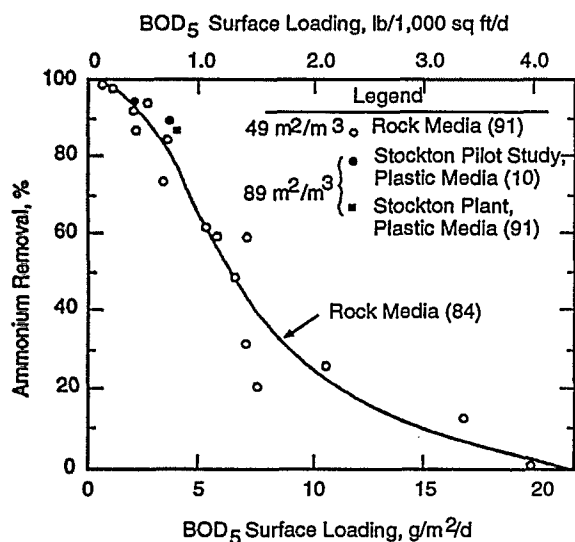

Figure 6-13. Effect of BOD5 surface loading on nitrification efficiency of rock and plastic media trickling filters (adapted from Reference 73).

The rock media BOD_5 volumetric loading rates reported at Stockton and for other rock (Figure 6-13) and slag media trickling filters are comparable; nearly complete nitrification is accomplished at BOD_5 loading rates in the range from 80 to 120 g/m³/d (5.0–7.5 lb/1,000 cu ft/d). The soluble BOD_5 loading rate of 1.8 g/m²/d (0.4 lb/1,000 sq ft/d) referenced earlier in Section 6.5.1 for nearly complete nitrification (effluent NH_4^+-N of approximately 2.0 mg/L) agrees with the Stockton results that on a total BOD_5 basis a surface loading of about 2.5 g/m²/d (0.51 lb/1,000 sq ft/d) is necessary for ≥90-percent ammonia nitrogen removal with rock and plastic media.

The BOD_5 volumetric and surface loading rates that are suggested for nearly complete nitrification are approximate. The actual loading rates that are required will be influenced by such factors as the recirculation ratio, the hydraulic surface loading rate, the filter depth, the specific hydraulic pattern and retention time in the trickling filter (when plastic media are employed), and environmental conditions. A more detailed discussion of the effects of such factors on the performance of trickling filters in general is presented elsewhere (49). Included is information concerning appropriate strategies to overcome issues common to trickling filters, whether they are designed for carbonaceous oxidation or carbonaceous oxidation and nitrification, such as biofilm sloughing cycles and the oc-

currence of filter flies and snails. A portion of the biofilm or biological slime will be sloughed either continuously or periodically during normal trickling filter operation. Accumulation of thick biofilms can result in oxygen limitations, which will impair performance; continuous and uniform sloughing, as measured by a relatively constant level of SS from the trickling filter, provides one indication of a well-operating unit.

Design Factors. Trickling filters are generally designed with at least a minimum effluent recycle capability to maintain a stable hydraulic loading during normal diurnal variations (87). The studies completed at Stockton on plastic media trickling filters (91) and those noted previously from operation of rock and slag media filters (1) indicate an increase in nitrification efficiency when recirculation is practiced. Investigators at Stockton noted an increase in nitrification performance from 67 to 87 percent when the recirculation ratio was increased from 2.7 to 3.8 and air circulation was increased by operating more ventilation fans. Both of these actions increased the DO concentration in the bulk liquid, resulting in performance improvement. Effluent recycling can also cause some degree of denitrification to occur in the upper portion of the trickling filter (73).

The use of effluent recirculation, the magnitude of the recirculation ratio and the recirculation configuration (e.g., before or after secondary clarification) are all factors affecting trickling filter performance. There is general agreement (49) that when consideration is given to both economics and performance for the typical tricking filter plant flowsheet (consisting of primary clarification, trickling filter(s), and secondary clarification), the effluent stream that is recycled should be the one after the trickling filter(s), and selection of a recirculation ratio of 1.0 is appropriate. Other hydraulic considerations include the trickling filter hydraulic loading, and the design and operation of the influent flow distributor. These considerations are common to the design of trickling filters for either carbonaceous oxidation or carbonaceous oxidation and nitrification and are dealt with in detail elsewhere (49). The value selected for the minimum hydraulic loading should ensure complete media wetting under all influent conditions. The value will depend on the characteristics of the media employed. Typical values are 1 to 3 $m^3/m^2/hr$ (0.41 to 1.22 gpm/sq ft) based on the cross-sectional or plan area of the filter. Results from one study (92) indicate that no advantage is gained from effluent recirculation as long as complete wetting is always accomplished.

The specific hydraulic pattern and retention time in plastic media trickling filters will influence the BOD_5 loading necessary to achieve nitrification. As previously noted, plastic media with cross-flow characteristics, when compared to vertical flow media, increase the hydraulic retention time or contact time between the biofilm and the influent and provide superior oxygen transfer (88). The results from

pilot studies completed at Garland, Texas (Figure 6-14), verify the claimed advantage for the cross-flow media, based on nitrification performance as a function of BOD_5 surface loading. The results were also compared to those from the operation of rock media trickling filters at Stockton, CA (91) and were found to be comparable (Figure 6-14). The results also indicated perhaps a slight efficiency improvement for 60° vs. 45° cross-flow media, and a significant efficiency advantage for the structured, corrugated plastic media over randomly packed plastic ring structures (73).
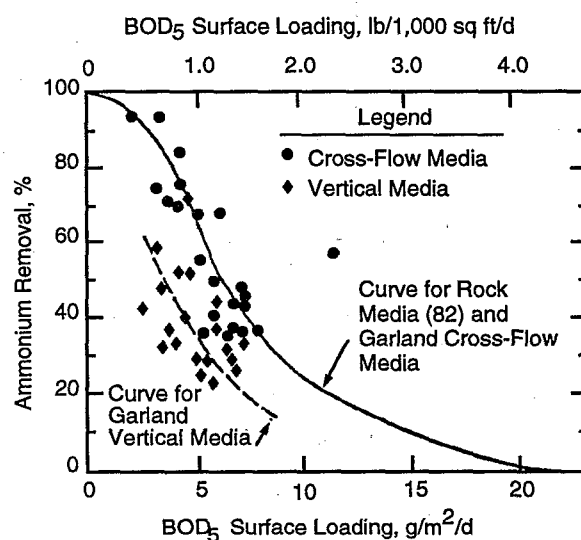

Figure 6-14. Comparison of performance of various trickling filter media (adapted from Reference 73).

Operating conditions that increase the DO concentration in the trickling filter liquid often improve performance, as previously noted. The effect of this environmental factor on performance is consistent with predictions from the biofilm model (Section 3.4). Biofilm modeling also indicates that pH conditions in the trickling filter liquid below a certain critical level may influence nitrification performance under certain conditions (93). Normally significant pH effects can be avoided by ensuring that the effluent alkalinity is equal to or greater than 50 mg/L, as $CaCO_3$ (93).

The importance of the DO concentration often masks the effect of pH (93) and temperature (94) on nitrification in trickling filters, particularly at a high carbonaceous feed concentration. In fact, a negative temperature correlation with TKN surface removal rate was shown by one author (87). Accounting for temperature effects during design is normally not necessary unless temperatures below 15°C are anticipated. At lower temperatures, pilot studies are recommended to evaluate the effect of temperature and its interaction with DO and NH_4^+-N, and other factors such

as hydraulic loading. Alternatively, temperature sensitivity relationships presented in Section 6.5.2.2 for trickling filters operating at low carbonaceous feed concentrations are acceptable, although using this approach can lead to a conservative design.

The importance of DO concentration in the operation of all trickling filters highlights the need for sufficient ventilation. If enough passageways are provided, the differences in air and wastewater temperatures and humidity differences between the ambient air and the air in the trickling filter provide a natural draft. This mechanism may provide the necessary aeration requirements on occasion, but not consistently (49). After summarizing the literature regarding ventilation rates in trickling filters it was recommended that for combined BOD_5 removal and nitrification systems the air flow requirements are given by (49,95):

$$\text{Air flow} = (150)(1.25 \; BOD_5 + 4.6 \; TKN)(PF)/1,440$$

where air flow is in m^3/min, and BOD_5 and TKN are in kg/d

Design Approach. Historically, engineers have selected an appropriate BOD_5 surface loading as a function of temperature to design trickling filters for nitrification of municipal wastewater at high $CBOD_5$ (87). As discussed earlier, a temperature correction is likely not warranted, at least at temperatures above 15°C (59°F). The use of a BOD_5 surface loading rate, specific to the media to be utilized in the biotower, remains an acceptable approach to biotower sizing for nitrification. More recently, one design manual has used a manufacturer's empirical correlation for calculating the average TKN surface removal rate in illustrating an approach to determine the biotower media requirements (49). TKN removal is due to the combined effects of biomass growth and nitrification; the correlation, as displayed on Figure 6-15, developed in concert with consulting engineers (49), was derived from an extensive array of operating results. Alternatively the design can be supported by pilot plant studies completed with the wastewater in question. Once a loading or removal rate has been selected and the media surface area requirements determined, the design procedure is basically in accord with that presented for the design of trickling filters at low feed $CBOD_5$ (Section 6.5.2.3).

Performance Information. In the assessment completed in 1991 (87), 2 of the 10 nitrifying trickling filter plants were identified as operating at a high carbonaceous feed concentration. These plants are located in Amherst, Ohio, and Wauconda, Illinois, respectively.

The wastewater treatment sequence at Amherst, Ohio, involves screening, grit removal, primary clarification, trickling filters, secondary clarification and chlorine disinfection (87). Two trickling filters are operated in series with no intermediate clarification. Each filter is 27.4 m (90

ft) long, 12.2 m (40 ft) wide, and 5.2 m (17 ft) deep and is packed with cross-flow plastic media.

Operating and performance information pertaining to the trickling filters over the 12-month period from February 1989 to January 1990 is presented in Table 6-18. During this period the wastewater flow was approximately 88 L/s (2 mgd), equivalent to its design flow. The hydraulic loading rate to the trickling filters averaged 23.0 $m^3/m^2/d$ (565 gpd/sq ft), and no effluent recirculation was practiced. The plant is required to meet an effluent ammonium-nitrogen limit of 3 mg/L during the summer months and 6 mg/L during the winter months. The raw wastewater temperature was less than 15°C (59°F) from February 1989 through May 1989, and again from November 1989 through January 1990, based on average monthly values (87). The average monthly effluent ammonium nitrogen values during these colder temperature periods ranged from 1.8 to 4.9 mg/L (Table 6-18). The BOD_5 media surface loading ranged from 1.17 $g/m^2/d$ (0.24 lb/1,000 sq ft/d) to 2.05 $g/m^2/d$ (0.42 lb/1,000 sq ft/d).

Over the entire period for which effluent NH_4^+-N data are available (i.e., April 1989 through January 1990), the average effluent ammonia-N concentration was 2.5 mg/L, the average raw wastewater temperature was 15.3°C (59.5°F), and the average BOD_5 surface loading was 1.36 $g/m^2/d$ (0.28 lb/1,000 sq ft/d). The Amherst results imply a temperature dependency for nitrification below 15°C (59°F). Operation and performance of the trickling filters would have to be assessed in more detail to determine whether the temperature effect implies the need for more trickling filter surface area or the need for modification of the operating conditions.

Contrary to the Amherst results, nitrification at the Wauconda, Illinois, facility appeared to be generally unaffected by temperature, based on operating and performance information for a two-year period beginning in January 1987 (87). The effect of temperature may have been masked by the low effluent ammonia values throughout the period. Two cross-flow plastic media trickling filters operate in parallel at Wauconda, receiving primary treated wastewater. Note that the Wauconda flowsheet includes an aerated solids contact flocculation step designed to improve solids capture during secondary clarification. First developed in 1979 (96), this trickling filter/solids contact (TF/SC) process has seen widespread use through the 1980s. In its simplest form the TF/SC process consists of a trickling filter, an aerated suspended growth contact reactor, a flocculation zone, and a secondary clarifier (Figure 6-16). Normally the contact reactor is not designed to nitrify (see Section 6.6.2.2 for further information).

Operating and performance information pertaining to the Wauconda trickling filters during 1988 is presented in Table 6-19. During this period the monthly average wastewater flow was 29 L/s (0.67 mgd), which is less than 50
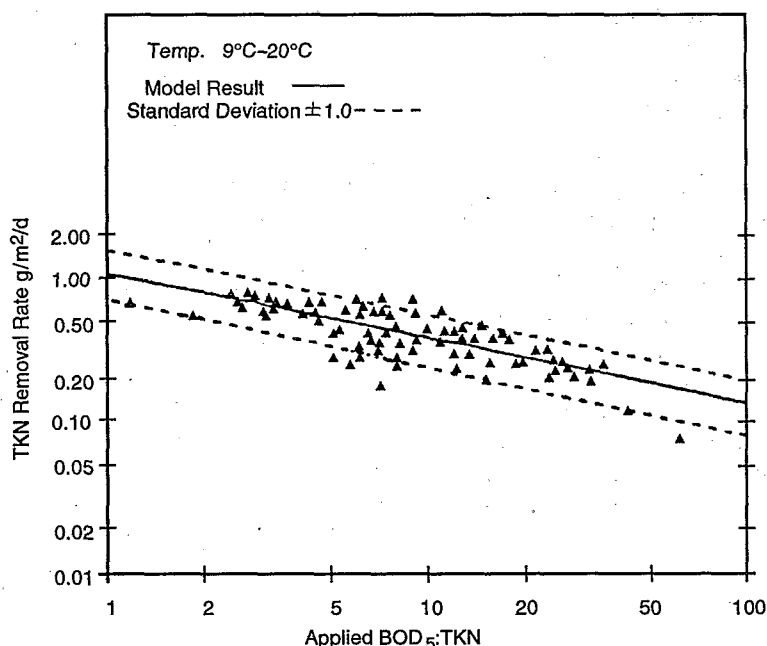
170

Figure 6-15. Correlation between TKN surface removal rate and wastewater BOD5:TKN proposed by American Surfpac Corporation (adapted from Reference 49).

percent of the plant's design flow. The hydraulic loading rate to the trickling filters was only 0.58 $m^3/m^2/hr$ (0.24 gpm/sq ft) and no effluent recirculation around the trickling filters was practiced. The plant is required to meet an effluent ammonia-nitrogen limit of 1.4 mg/L during the summer months and 4 mg/L during the winter months. In 1988, the raw wastewater temperature was less than 15°C (59°F) from January through May, based on average monthly values (87). The average monthly secondary effluent ammonia-nitrogen was 0.24 mg/L and the BOD_5 media surface loading was 1.92 $g/m^2/d$ (0.39 lb/1,000 sq ft/d) during this same five-month period (Table 6-19). The performance of the Wauconda plant may have been influenced by nitrification occurring in the contact reactor, as a result of the low flow condition and the introduction of sloughed trickling filter nitrifiers.

In the assessment completed in 1991 (87), detailed performance information for periods of 12 months or more was reported for five nitrifying trickling filter plants operating at a high carbonaceous feed concentration. The effluent ammonia nitrogen as a function of BOD_5 surface loading for these plants is plotted in Figure 6-17, based on the mean of the reported average monthly values. The Wauconda and Ashland plants both incorporate a solids contact step prior to secondary clarification. These results imply that BOD_5 media surface loading rates below 1.0 $g/m^2/d$ (0.2 lb/1,000 sq ft/d) may be best to ensure residual ammonium-nitrogen concentrations less than 2.0 mg/L.

In summary, the following process factors are some that affect the design of trickling filters for nitrification at high carbonaceous feed concentrations:

- Hydraulic loading

- Hydraulic pattern and retention time in the filter media

- DO concentration in the filter liquid

- pH and temperature of the filter liquid under certain conditions

- Feed TKN concentration

- Feed BOD_5 concentration

6.5.2.2 Nitrification Design and Performance Information at Low Carbonaceous Feed Concentration

Many operating trickling filters in the United States are designed to treat municipal wastewaters that receive a low carbonaceous feed concentration. Many of these trickling filters are the second stage of a two-stage trickling filter plant with intermediate clarification. When operated in this fashion, the trickling filter is often referred to as a separate-stage or tertiary trickling filter. Tertiary trickling filter nitrification has been classified in one municipal wastewater treatment design manual (49) by conditions wherein the influent wastewater BOD5:TKN is less than 1.0 and the $SBOD_5$ is less than 12 mg/L.

In general, low levels of organics in the influent to attached growth reactors can be advantageous. Ideally, such conditions result in low effluent SS, potentially elimi-

171

Table 6-18. Amherst, Ohio, Wastewater Treatment Plant Carbonaceous Oxidation-Nitrification System Operating Conditions and Performance Results, February 1989 to January 1990 (Adapted from Reference 87)*

Month	Waste-water Flow L/s (mgd)	BOD₅, mg/L Primary Effluent	BOD₅, mg/L Secondary Effluent	TSS, mg/L Primary Effluent	TSS, mg/L Secondary Effluent	Ammonia-N, mg/L Primary Effluent	Ammonia-N, mg/L Secondary Effluent	Trickling Filter BOD₅ Loading g/m²/d (lb/1,000 sq ft/d)	Trickling Filter BOD₅ Loading g/m³/d (lb/1,000 cu ft/d)
Feb.	91 (2.06)	89	12	74	12	12.2	—	2.05 (0.42)	200 (12.5)
Mar.	106 (2.42)	62	11	56	12	14.3	—	1.66 (0.34)	163 (10.2)
Apr.	120 (2.74)	62	8	88	9	9.3	3.6	1.86 (0.38)	184 (11.5)
May	106 (2.42)	62	10	118	15	7.6	2.2	1.66 (0.34)	162 (10.1)
June	102 (2.32)	40	8	216	10	8.9	2.0	1.03 (0.21)	101 (6.3)
July	90 (2.04)	61	8	196	9	14.3	1.2	1.37 (0.28)	136 (8.5)
Aug.	69 (1.57)	80	7	192	8	18.4	1.6	1.37 (0.28)	136 (8.5)
Sept.	67 (1.52)	67	6	183	9	17.1	1.9	1.12 (0.23)	112 (7.0)
Oct.	67 (1.52)	56	5	132	7	17.0	1.4	0.93 (0.19)	93 (5.8)
Nov.	78 (1.78)	59	4	99	8	17.8	1.8	1.17 (0.24)	114 (7.1)
Dec.	77 (1.75)	99	6	129	9	—	4.9	1.91 (0.39)	189 (11.8)
Jan.	101 (2.29)	48	6	89	10	—	4.3	1.22 (0.25)	120 (7.5)

* Average monthly values are presented. Primary effluent values are calculated assuming 35-percent removal of BOD₅ and TSS in primary clarifier. The primary effluent is the feed to the trickling filters. Trickling filter surface loading is based on total media surface area of the two trickling filters operating in series.
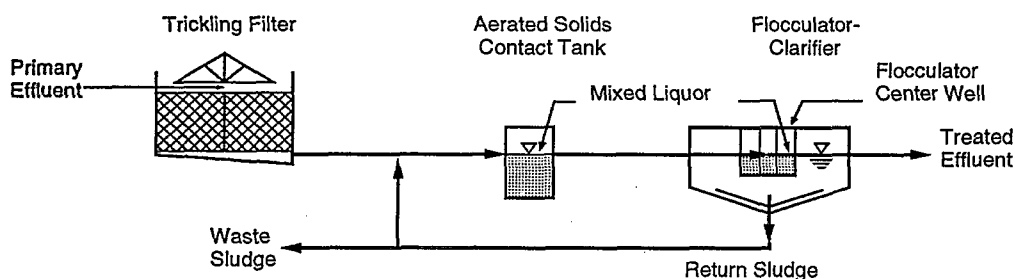

Figure 6-16. Common process variation of the trickling filter solids contact process (adapted from Reference 96).

Table 6-19. Wauconda, Illinois, Wastewater Treatment Plant Carbonaceous Oxidation-Nitrification System Operating Conditions and Performance Results, January 1989 to December 1989 (Adapted from Reference 87)*

Month	Wastewater Flow L/s (mgd)	BOD_5, mg/L Primary Effluent	BOD_5, mg/L Secondary Effluent	TSS, mg/L Primary Effluent	TSS, mg/L Secondary Effluent	Ammonia-N, mg/L Primary Effluent	Ammonia-N, mg/L Secondary Effluent	Trickling Filter BOD_5 Loading $g/m^2/d$ (lb/ 1,000 sq ft/d)	Trickling Filter BOD_5 Loading $g/m^3/d$ (lb/ 1,000 cu ft/d)
Jan.	44 (0.99)	78	15	70	10	12.4	0.4	1.90 (0.39)	188 (11.71)
Feb.	36 (0.83)	77	9	125	4	17.6	0.3	1.56 (0.32)	155 (9.69)
Mar.	35 (0.80)	101	11	79	14	17.0	0.3	2.00 (0.41)	196 (12.26)
Apr.	44 (1.00)	76	9	104	20	14.4	0.1	1.86 (0.38)	185 (11.53)
May	28 (0.64)	146	10	81	8	14.1	0.1	2.30 (0.47)	227 (14.17)
June	24 (0.55)	128	8	73	9	15.6	0.2	1.76 (0.36)	171 (10.68)
July	24 (0.55)	123	8	59	4	14.3	0.1	1.66 (0.34)	164 (10.26)
Aug.	27 (0.61)	124	10	68	7	13.3	0.3	1.86 (0.38)	184 (11.47)
Sept.	24 (0.55)	109	9	75	9	16.2	0.3	1.47 (0.30)	146 (9.09)
Oct.	21 (0.48)	122	9	81	6	17.0	3.6	1.47 (0.30)	142 (8.88)
Nov.	26 (0.60)	104	15	60	17	14.6	0.7	1.56 (0.32)	152 (9.47)
Dec.	21 (0.48)	115	12	61	22	13.1	0.3	1.37 (0.28)	134 (8.37)

* Average monthly values are presented. The primary effluent is the feed to the trickling filters. Secondary effluent is after solids contact/flocculation clarification steps. Trickling filter surface loading is based on media surface area.
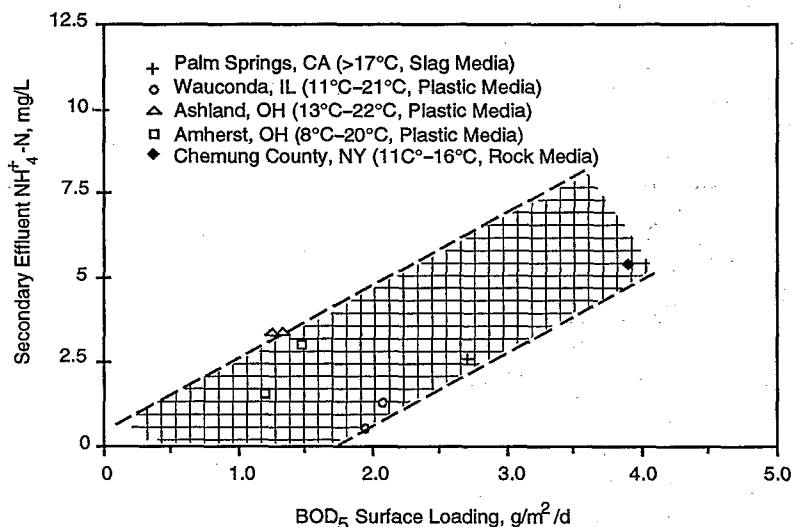

Figure 6-17. Effect of BOD_5 surface loading on nitrification performance (adapted from Reference 87).

nating the need for a downstream solids-liquid separation step (as discussed in Section 6.3.1). Low levels of organics in the influent result in little biomass production, which in turn leads to reduced biofilm formation. Pluggage of voids in the media and resulting bypass or short circuiting flow patterns are of less concern than when operating at a higher carbonaceous feed concentration.

Nitrification in plastic media trickling filters operating at low $CBOD_5$ was first reported in the period from 1973 to 1975 and the results are incorporated into the first EPA Process Design Manual for Nitrogen Control (1). The results are expressed in the form of design curves which specified the surface area required per mass of ammonium oxidized per day as a function of effluent ammonium. Although lacking a fundamental basis, these curves were used for design for several years after the Manual was introduced in 1975. Since that time, research and development efforts have improved our understanding of how various factors affect the performance of the trickling filter when operated at a low $CBOD_5$ concentration.

By operating at a low $CBOD_5$, the influence of bulk liquid DO and ammonium-nitrogen concentrations, environmental factors (i.e., temperature and pH), and other operating and design conditions (e.g., hydraulic loading rate and media configuration) on the nitrification process can be better understood. Nitrifiers are dominant because of the reduced competition with heterotrophic growth for DO and other nutrients. Still, previously mentioned (Section 6.5.1) factors such as nonhomogeneous biofilm thickness and density will generally prevent direct application of biofilm models for design purposes.

Design Factors. The rate of nitrification in attached growth reactors, in the absence of significant $CBOD_5$, is influenced by the concentration of both ammonium and DO in the liquid phase. Depending on their concentration values, either substrate can dictate the oxidation rate of ammonium per unit surface area. The nitrification rate no longer is zero-order with respect to ammonium concentration when the concentration reaches a critical value. Although somewhat controversial, this value appears to be in the range from 2 to 5 mg/L NH_4^+-N (84,97). Ammonium surface removal rates may appear diminished at lower ammonium concentrations; the reason for the lower rate, however, may be ineffective surface area due to poor biofilm coverage at the location in the filter where the ammonium concentration is the lowest (84). This apparent reduction in rate still implies the need for more media surface area, or specific operating procedures that will overcome the issue of incomplete biofilm surface coverage.

At high ammonium concentrations (i.e., greater than 5 mg/L NH_4^+-N), nitrification is limited by oxygen mass transfer across the liquid film and not by the DO concentration in the bulk liquid phase, according to some researchers (98). These results imply that oxygen availability in certain regions of the trickling filter appears to regulate the nitrification rate. Data compiled from five separate studies (97) indicate that as the ammonium-nitrogen load is increased beyond a critical value, the corresponding removal rate is not predictable (Figure 6-18). The reasons for the scatter in the data above a critical ammonium-nitrogen loading value of approximately 1 $g/m^2/d$ (0.2 lb/sq ft/d) (Figure 6-18) could be due to differences in oxygen transfer as a result of different media configurations (i.e., vertical versus cross-flow), or due to differences in environmental and operating conditions at the different sites where the studies were completed. The scatter also could be due to the problem of using average rates when correlating the data (84). The authors attempted to estimate bulk liquid DO conditions for the different data sets while separating out the effects of media configuration, operating conditions, and environmental factors (97). Figure 6-19 is typical of the results of their analysis, where one data set is presented from operations using vertical media at constant hydraulic load and recycle conditions, and at one temperature condition. The results (Figure 6-19) imply a constant correlation between ammonium-nitrogen surface loading and removal rate for a given bulk liquid DO concentration.

Based on the previous discussion of the effects of DO and NH_4^+-N concentration on ammonium removal rates, the NH_4^+-N profile in a trickling filter operating at low influent $CBOD_5$ can be predicted. The typical profile will be a straight line reduction of NH_4^+-N at a removal rate controlled by oxygen availability in the upper portions of the filter. The removal rate will decrease as the rate-limiting factor becomes ammonium rather than oxygen diffusion. The inflection point, although dictated by oxygen mass transfer, temperature, and degree of biofilm coverage on the media, will typically occur when the ammonium-nitrogen concentration is in the range from 2 to 5 mg/L. Bulk DO concentrations that are less than 70-percent saturation (i.e., approximately 6 to 7 mg/L) reflect inadequate trickling filter ventilation (98).

The practice of storage and control of ammonium-laden anaerobic digester supernatant return streams to the trickling filter has been proposed as a means of leveling out the ammonium load to the filter during a 24-hour period (84). This practice should ensure more consistent ammonium penetration to the lower filter depths, ideally eliminating patchy biofilm coverage. A related biofilm control procedure proposes regular flooding and backwashing to eliminate the growth of certain predators such as filter fly larvae, which may cause a decrease in nitrification rates, and to prevent repetitive biofilm sloughing (84). Such procedures may eliminate the need for post clarification or filtration, depending on effluent quality requirements (84). Others have proposed the use of high-intensity flushing to control the growth and development of filter flies and, possibly, snails (49).
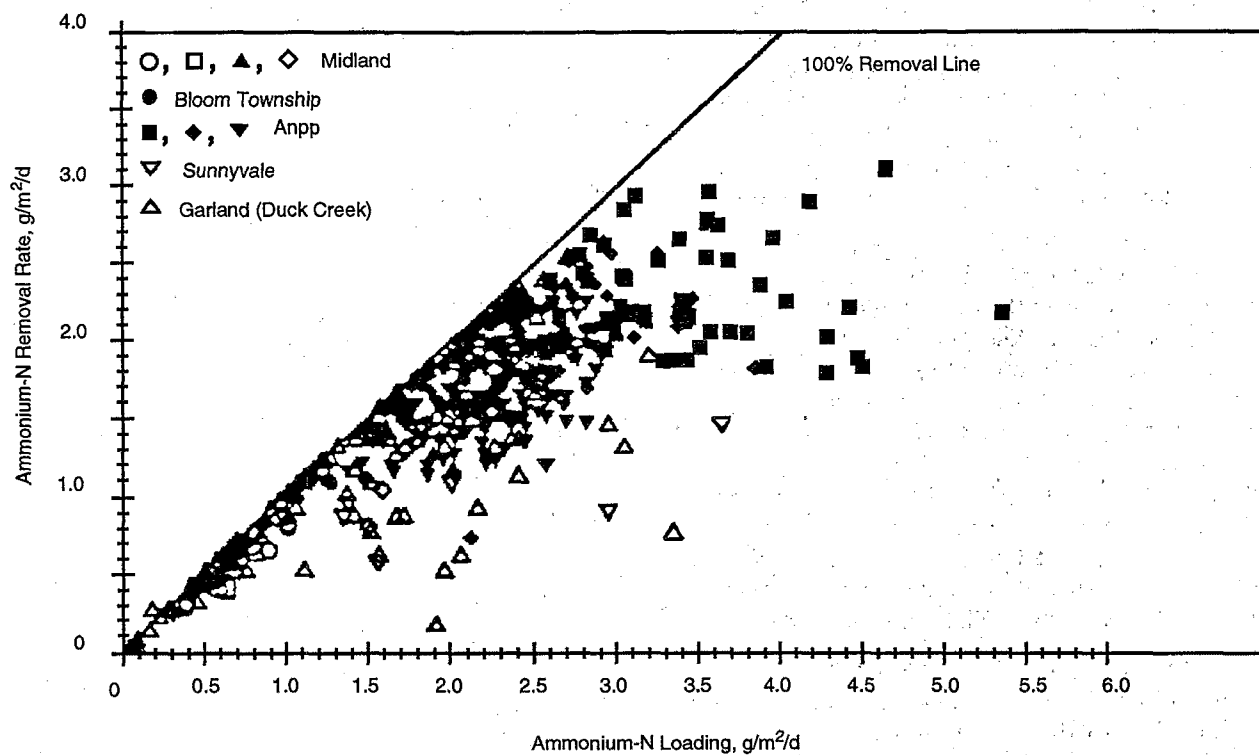
174

Figure 6-18. Ammonium surface loading versus removal rate (adapted from Reference 97).
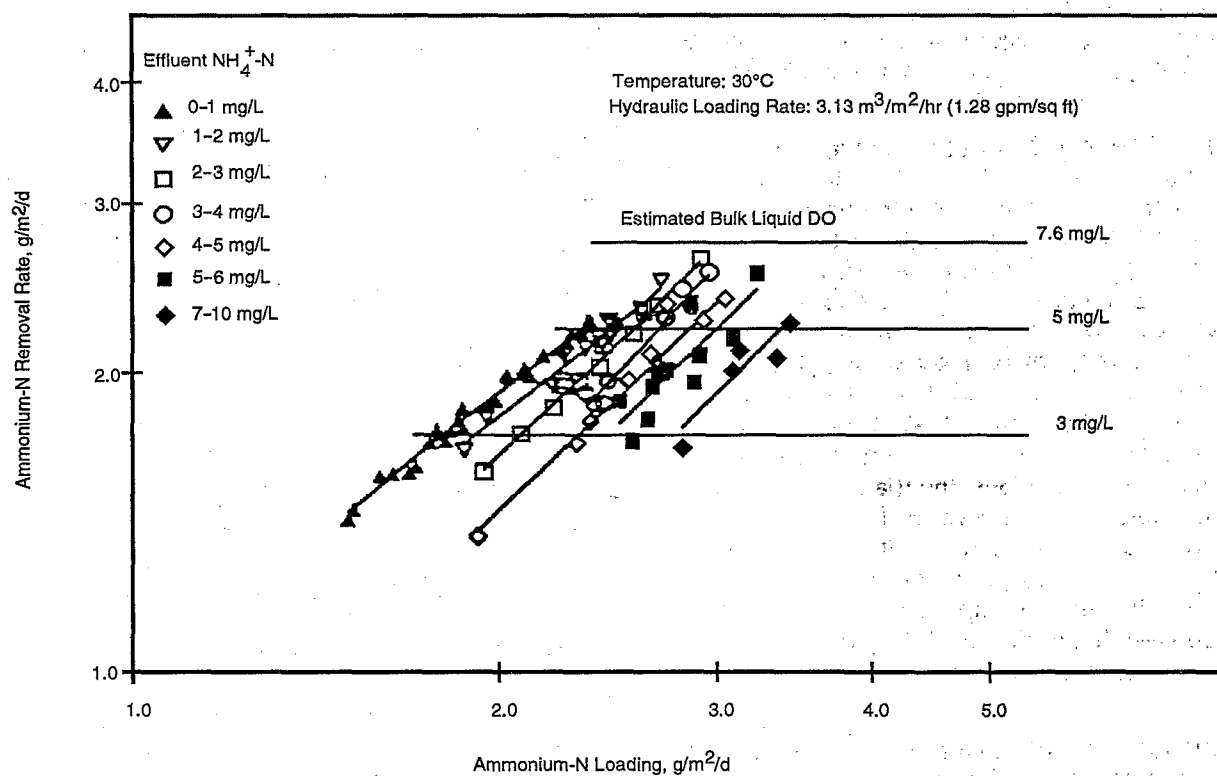

Figure 6-19. Effect of bulk liquid DO conditions on surface loading versus removal rate correlation (adapted from Reference 97).

The effect of temperature on the nitrification rate in trickling filters operating at low CBOD$_5$ will be influenced by such factors as oxygen availability, influent and effluent ammonium concentration, and hydraulic loading conditions; thus the reported effect of temperature is variable (85,99). Zero-order or maximum rates of nitrification appear to be more influenced by oxygen mass transfer characteristics of the media (84), which in turn are influenced only moderately by temperature as opposed to intrinsic nitrifier growth rates. These observations are consistent with the biofilm model previously noted (Section 6.5.1). A comparison of the effect of temperature on zero-order rates versus rates limited by various factors is presented in Figure 6-20. The higher nitrification rates at Central Valley are attributed partially to the use of cross-flow media, which has oxygen transfer characteristics superior to those at Midland and Lima, where vertical flow media are used. The biofilm control procedures adopted in the Central Valley pilot study were also credited for the higher nitrification rates. The poor performance at Zurich was attributed to poor wetting and biofilm coverage (84).

When nitrifying trickling filters, which are operating at low influent CBOD$_5$ and near neutral pH conditions, are temporarily exposed to a lower pH, a significant short-term (less then two days) negative effect on the nitrification rate will be observed, according to Swiss researchers (93). Based on other scientists' experience (100), these researchers concluded that the pH dependence is likely
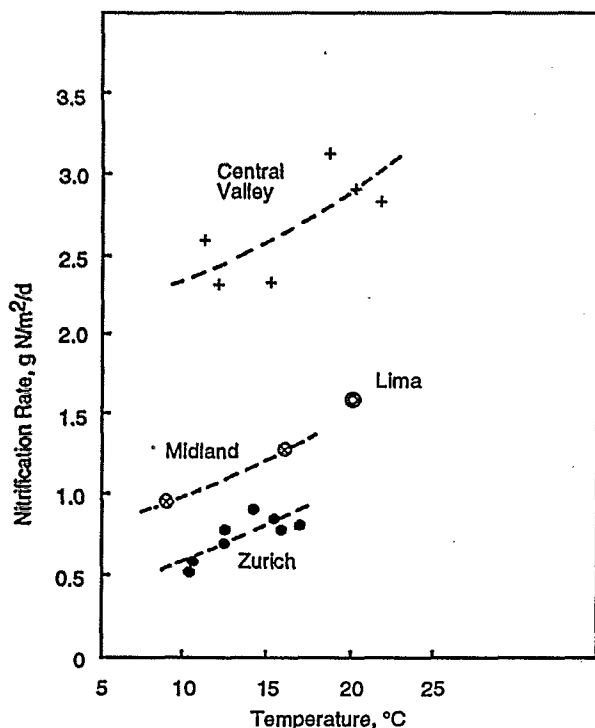
only temporary. The Swiss findings are particularly relevant to nitrifying trickling filters exposed to significant variations in influent ammonium concentrations when operating at low effluent alkalinity values (less than 50 mg/L as CaCO$_3$) (93).

Although authors have presented data indicating that a higher hydraulic loading increases the nitrification rate (97,99), the hydraulic loading required to maximize the rate is unknown (49). The data compiled from the five separate studies discussed earlier indicate that a hydraulic loading above 3.6 m^3/m^2/hr (1.47 gpm/sq ft) produced the highest ammonium-nitrogen surface removal rates (97). Interpretation of the data is complicated by how the interaction between hydraulic loading and oxygen availability affects the rates.

The characteristics of the medium employed will dictate the minimum hydraulic loading needed both to ensure complete wetting and to affect the ammonium-nitrogen volumetric removal rate. The superior oxygen transfer characteristics and higher specific surface area of cross-flow media favor its use over vertical flow media for nitrification of wastewaters containing a low influent CBOD$_5$.

One author recommended the use of medium density, cross-flow media (specific surface area of 138 m^2/m^3, or 42 sq ft/cu ft) over a higher density alternative (specific surface area of 223 m^2/m^3, or 68 sq ft/cu ft), citing issues of poorer wetting characteristics and a greater tendency for plugging associated with the higher density material (84). Less efficient wetting characteristics of the higher density media result in a lower ammonium-nitrogen surface removal rate, since a portion of the surface area is ineffective. Swiss researchers recommend the selection of a media with a specific surface area in the range from 150 to 200 m^2/m^3 (46 to 61 sq ft/cu ft) (101). High density media trickling filters have proved to be effective at Reno-Sparks, Nevada (102). The difference may be explained by the Reno-Sparks investigators' observation that grazing organisms were not a problem, as they were for the Swiss investigators. Problems with grazing organisms at pilot scale with fixed distributors may not be realized at full scale with rotary distributors (103).

The fact that plugging is often associated with higher density media implies the need for low CBOD$_5$ and TSS concentrations in the biotower influent. The role of CBOD$_5$ in reducing ammonium-nitrogen surface removal rates as a consequence of heterotrophic growth competition, has been discussed. Extensive biomass growth also can affect the hydraulic integrity of the biotower causing influent short-circuiting or media plugging, or requiring flow bypassing. Researchers have found that particulate CBOD$_5$ affects nitrification rates to the same negative degree as soluble CBOD$_5$ (104).

Figure 6-20. Effect of temperature on nitrification rates in trickling filter systems (adapted from Reference 96).

Design Approach. The preceding discussions imply that trickling filters operating with low feed CBOD$_5$ should be

designed to maximize the fraction of the media volume operating at the maximum nitrification rate, or the rate that is not limited by the concentration of NH_4^+-N. Furthermore, the design should provide the maximum hydraulic loading possible without the need for recirculation and should provide adequate air flow to maximize oxygen transfer to the liquid.

Given these considerations, trickling filters should be constructed as deep towers with lower cross-sectional areas, and should be designed with forced air ventilation. Nitrification biotowers have been constructed with depths as great as 12.8 m (42 ft); shallower towers can be employed by operating two units in series. It has been demonstrated that if biotowers are designed in series, better performance can be expected, as long as the order of operation of the two units can be reversed every few days (101). This operating protocol will lead to development of a more uniform biofilm in the two towers. Ensuring air flow throughout the trickling filter requires excess air distribution within the filter (49). It has been suggested that during normal and low NH_4^+-N loading periods, the minimum air flow rate should supply 50 kg O_2/kg of O_2 required (49). During peak loading periods a value of 20 to 30 kg O_2 supplied per kg O_2 required is claimed to be acceptable (49).

The procedure used to design nitrifying trickling filters at low feed $CBOD_5$ requires specification of the following parameters:

1. The full range of NH_4^+-N surface removal rates across the biotower under the anticipated temperature conditions.

2. The type of media and its oxygen transfer characteristics.

3. The minimum biotower hydraulic loading rate and the corresponding maximum tower surface area.

4. Any minimum or maximum biotower depth limitations.

5. The oxygen requirement and the minimum air flow necessary to meet that requirement.

6. The design of the influent distributor.

Approaches to determine the required design parameters have been proposed by researchers, consultants, and system suppliers (49). One approach (84,96), supported by theoretical considerations and pilot plant results, is based on the efforts of Swiss researchers (70) and others (92). The approach involves the use of the following equation to determine the NH_4^+-N surface removal rate at any point in the biotower:

$$J_N(Z,T) = J_{N\,max}\left[\frac{N}{K_N + N}\right]e^{-kz} \qquad (6\text{-}26)$$

where:

J_N (Z,T) = NH_4^+-N surface removal or nitrification rate at depth Z and temperature T, g/m²/d
Z = depth in tower, m
T = temperature, °C
$J_{N\,max}$ = maximum NH_4^+-N surface removal rate at temperature T, g/m²/d
N = bulk liquid NH_4^+-N concentration, mg/L
K_N = saturation constant, mg/L
k = empirical parameter describing decrease in rate with depth

$J_{N\,max}$ will be dictated by the oxygen transfer efficiency of the media employed according to:

$$J_{N\,max} = \frac{E(J_{O_2\,max})}{4.3} \qquad (6\text{-}27)$$

where:

E = media effectiveness factor
$J_{O_2\,max}$ = maximum surface oxygen transfer rate, g/m²/d

The factor 4.3 in Equation 6-27 represents the mass of oxygen consumed per unit mass of ammonium-nitrogen oxidized. $J_{O2\,max}$ is calculated from the Logan model (92). Departure of E from a value of 1.0 reflects incomplete media wetting, impact of predators on the biofilm, or competition from heterotrophs for oxygen (84). The term $\frac{N}{K_N + N}$ in Equation 6-26 is equivalent to a Monod saturation term and was used by the Swiss researchers to describe the influence of ammonium concentration on the nitrification rate (70). The term e^{-kz} was used to describe the nitrification rate decline observed with depth and attributed to poor biofilm growth.

Two solutions to Equation 6-26 have been presented as follows (70):

$$\frac{a\,J_{N\,max}}{K\,v_H}(1 - e^{-kz}) = N_i - N + K_N \ln\left(\frac{N_i}{N}\right) \qquad (6\text{-}28)$$

or for the case where k=o:

$$\frac{z\,a\,J_{N\,max}}{v_h} = N_i - N + K_N \ln\left(\frac{N_i}{N}\right) \qquad (6\text{-}29)$$

where:

a = specific surface area of trickling filter media, m²/m³
v_h = hydraulic loading of trickling filter media (flow/cross-sectional area), m³/m²/d
N_i = ammonia-nitrogen concentration applied to the filter, g/m³

Where recirculation is used to obtain the desired hydraulic loading on the support media, the recirculation ratio, r, alters the value of ammonia nitrogen applied to the filter. The relationship is:

$$N_i = \frac{N_o + rN}{1 + r} \qquad (6\text{-}30)$$

Table 6-20. Calculated Trickling Filter Nitrification Model Parameters from Pilot Plant Studies (Adapted from Reference 84)

Period	Test Number	$J_{N\ max}$ g N/m³/d	lb N/1,000 sq ft/d	k, m⁻¹ (ft⁻¹)	E	Temp., °C
04/17/87–05/21/87	1	2.1	0.43	0	0.66	15.5
06/19/87–07/23/87	2	2.9	0.59	0.075 (0.023)	0.89	20.0
07/31/87–09/03/87	3	2.8	0.56	0	0.83	21.5
10/23/87–11/19/87	4	3.2	0.65	0.16 (0.05)	0.99	18
11/25/87–12/10/87	5	2.3	0.46	0	0.71	15
12/21/87–01/07/88	6	2.3	0.47	0	0.72	12
01/01/88–02/25/88	7	2.6	0.53	0	0.81	11

Table 6-21. Comparison of Measured and Predicted Nitrification Rates (Adapted from Reference 84)

Plant	Reference	Media Code	Period of Designation	Temp., °C	Apparent Zero-Order Nitrification Rate, g N/m²/d (lb N/1,000 sq ft/d)		Predicted Maximum Rate* g N/m²/d (lb N/1,000 sq ft/d)		E
Midland, MI	97,98	VFc27	Run 9A	13	1.2	(0.24)	1.4	(0.28)	0.86
			Run 12	7	0.93	(0.19)	1.3	(0.26)	0.74
Lima, OH	99	VFc27	TP2	18	1.2	(0.25)	1.4	(0.28)	0.88
			TP5	21	1.8	(0.37)	1.4	(0.28)	1.30
			TP6	22	1.5	(0.31)	1.4	(0.28)	1.10
			TP7	22	1.2	(0.24)	1.5	(0.32)	0.76
Bloom Twp., IL	100	VFc27	9/22–10/10	20	1.2	(0.24)	1.3	(0.27)	0.88
			10/11–10/20	17	1.1	(0.22)	1.3	(0.27)	0.82
Zurich, Switz. (3.9 m tower)	100	VFd28	N/A	17–20	1.6	(0.32)	1.7	(0.35)	0.89
		XFa68	N/A	17–20	1.2	(0.24)	3.1	(0.63)	0.38
Zurich, Switz. (6.8 m tower)	93	XFa68	N/A	13	1.1	(0.23)	2.8	(0.58)	0.3

* Predicted from Equations 6-26 and 6-27.

If recirculation is used, an iterative solution is necessary because recycle effects are included in both the N_i and v_h terms (84).

Values for the terms in Equations 6-26 and 6-27 were developed through profile sampling of a pilot trickling filter during studies at the Utah Central Valley Water Reclamation Facility (84,96). These are summarized on Table 6-20.

The value of K_N averaged between 1 and 2 mg/L during the pilot study (84). The $J_{N\ max}$ rates presented in Table 6-20 are the same rates presented in Figure 6-20 for Central Valley. The value of k in Table 6-15 varied from 0, implying no rate decrease with media depth, to 0.16. The range in the value of k is significant.

The nitrification design model was used to analyze certain data collected from previous studies (84). The data se-

lected (Table 6-21) were from periods when the biotowers appeared to be operating in the zero-order region throughout their depths, as evidenced by effluent ammonium levels of 3.0 to 5.0 mg/L. The measured rates thus represent the apparent maximum or zero-order nitrification rate. The predicted maximum rates were calculated from Equations 6-26 and 6-27, using the Logan model (92) to calculate $J_{O2\ max}$ for the specific corrugated sheet media employed. The measured and predicted rates show reasonable agreement for the vertical flow media data (Table 6-21, VF_c and VF_d media), reflected by values of E near 1.0. The low E values for the cross-flow media (Table 6-21, XF_a media) are believed by the authors to be caused by poor media wetting. The medium employed in the Zurich studies was the high density material (specific surface area of 223 m²/m³, or [68 sq ft/cu ft]) cited previously.

The vertical flow media nitrification rates are considerably lower than those reported from the Central Valley studies (Table 6-20). The medium employed at Central Valley was medium density, cross-flow material. The differences in the media, together with other operating factors, likely account for the difference in rates. In addition to selection of an ideal media, several other factors are proposed by the authors as being important to attaining these high rates (84):

- High quality secondary effluent (i.e., average TSS and $CBOD_5$ less than 15 mg/L) as influent to the biotower.

- Provision of flooding capability for prevention of filter fly larvae growth.

Table 6-22. Annual Operating Information from Five Nitrifying Trickling Filters Receiving Low Influent CBOD (Adapted from Reference 96)

Plant*	Influent Wastewater Flow, L/S (mgd)	Nitrification Rate, g N/m²/d (lb N/1,000 sq ft/d)	Temp., °C
A	572 (13.01)	0.29 (0.059)	14–24
B	70 (1.60)	0.29 (0.059)	9–19
C	70 (1.60)	0.20 (0.041)	14–20
D	582 (13.24)	0.39 (0.080)	10–23
E	823 (18.72)	0.34 (0.070)	12–27

* Plants A, B, and D are located in the Midwest. Plants C and E are located in the West.

- Use of forced ventilation to prevent stagnation.

- Use of any ammonium-laden supernatant return flows to even out ammonium loads.

- Use of regular backwashings for removal of SS accumulations; the need for tertiary clarification also may be eliminated.

Performance Information. Recent surveys identify approximately 20 operating trickling filters in 1990 that were designed for nitrification of municipal wastewater at low influent $CBOD_5$ (87,96). These trickling filters served as the tertiary treatment step for trickling filters or activated sludge systems designed for secondary treatment. Detailed operating and performance information suitable for evaluation was obtained from five of these plants (96). Annual operating information for the five plants is presented in Table 6-22. Performance information, in the form of monthly average effluent ammonia-nitrogen concentration plotted against the cumulative frequency, is presented in Figures 6-21 and 6-22. The results indicate that all plants were achieving less than 2.0 mg/L ammonianitrogen 50 percent of the time, with four of the plants at less than 2.0 mg/L 90 percent of the time. The plants
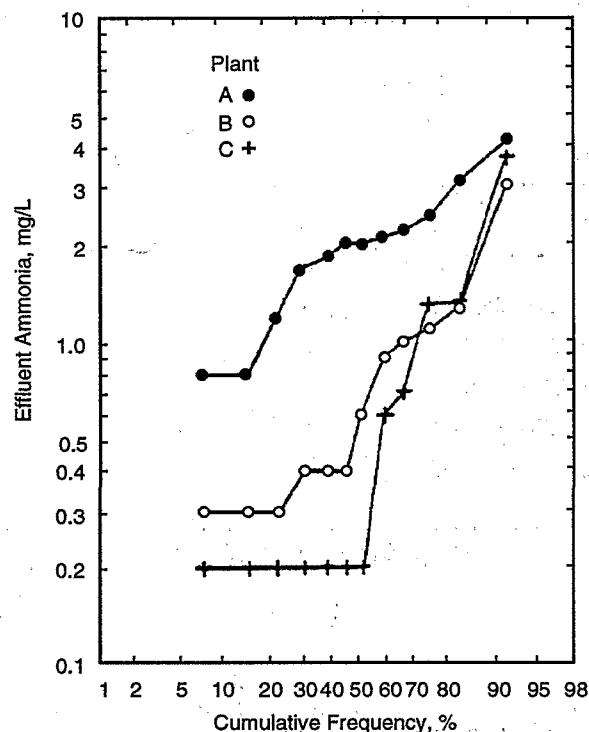

Figure 6-21. Nitrification performance on trickling filters receiving low enfluent $CBOD_5$—Plants A, B, and C (adapted from Reference 96).
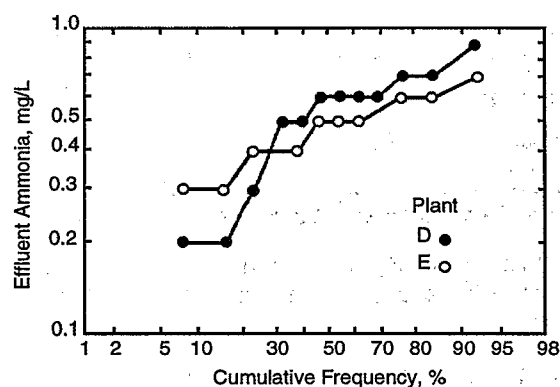

Figure 6-22. Nitrification performance of trickling filters receiving low influent $CBOD_5$—Plants D and E (adapted from Reference 96).

were all operating at somewhat conservative ammonium surface loading rates. Utilization of conservative practices in the design of tertiary nitrifying trickling filters to date is also reflected by the use of a solids-liquid separation step following the biotower (87,96). Approximately 90 percent of the plants employ either effluent filtration or clarification.

179

6.5.2.3 Design Example No. 3: Nitrification in a Trickling Filter at Low Carbonaceous Feed Concentration

This design example is based on information provided in Section 2.9 for the more complex Plant B. The calculations address sizing requirements to meet the less stringent effluent limits provided in Table 2-10 (with no limit for total N). A trickling filter is to be added to treat the existing secondary effluent. The plant flows and unsettled secondary effluent quality correspond to the values presented in Table 6-4. As in Design Example No. 2, assume that 21 mg/L of ammonia nitrogen are available for nitrification. For purposes of this example, assume also that the existing secondary effluent has the characteristics summarized in Table 6-23. Additional design information used for this example is also summarized in Table 6-23.

Table 6-23. Design Information for Nitrification of a Low Carbonaceous Feed Concentration and Less Stringent Effluent Requirements

Wastewater Flow Characteristics, m^3/d (mgd)	
Raw wastewater average flow	18,925 (5.0)
Total secondary effluent average flow	21,055 (5.5)
Actual Secondary Effluent Concentrations, mg/L	
Soluble COD	27
Nitrogen available for nitrification	21
Alkalinity as $CaCO_3$	120
Trickling Filter Reactor Effluent Characteristics, mg/L	
Soluble COD	20
Ammonia nitrogen	1.5
Design Conditions/Assumptions	
Reactor temperature, °C	15
Reactor pH range	7.0–7.6
Air flow rate (at average secondary loading)	
kg O_2 supplied/kg O_2 required	50

The process design steps are as follows:

1. Determine the process design factor to be used in sizing the trickling filter. Following from Chapter 2 (Table 2-18), the peaking factor developed to account for influent loading variations for attached growth systems is 1.92. However, for this situation assume that flow buffering due to upstream unit operations and effluent recirculation, the effluent quality requirement, anticipated variations in process conditions, and the uncertainty in the kinetic approach warrant an overall process design factor of 1.5. In sizing a trickling filter, this determination is equivalent to specifying a design nitrification rate ($J_{N\ des}$) equal to the maximum nitrification rate divided by the process design factor.

2. Verify that the trickling filter reactor operating pH can be expected to be in the assumed range of 7.0–7.6 (Table 6-23). Alkalinity destroyed can be calculated from the coefficient presented in Table 3-1 as:

(7.1 mg $CaCO_3$/mg N oxidized)(21 mg/L oxidizable N) = 149 mg/L as $CaCO_3$

Alkalinity available (Table 6-23) is 120 mg/L (as $CaCO_3$). It is reasonable to assume that the pH will remain in the required range since forced air ventilation is employed and a minimum residual alkalinity of 50 mg/L (as $CaCO_3$) is maintained. On this basis, the required supplemental alkalinity (ignoring any residual NH_4^+-N) will be:

(149 mg/L destroyed + 50 mg/L minimum residual) − 120 mg/L available = 79 mg/L (as $CaCO_3$) to be supplemented.

The mass of alkalinity required under average and maximum day conditions can be determined following the approach presented in Design Example No. 1 (Section 6.4.2.2).

3. Select the trickling filter media and establish the maximum NH_4^+-N surface removal rate $J_{N\,max}$. Select a medium density cross-flow media with a specific surface area of 138 m^2/m^3 (42 sq ft/cu ft). Assume biotower effluent recirculation is available to maintain a total influent to the biotower equivalent to the total secondary effluent average flow. Select an average flow hydraulic loading to the media sufficient to provide adequate wetting. For this example choose a hydraulic loading rate of 80 m/d. Based on pilot or full-scale operating information and application of Equations 6-26 and 6-27, assume $J_{N\,max}$ is 2.0 $g/m^2/d$ and k equals 0 when K_n is 1.5 mg/L.

4. Calculate a maximum design nitrification rate based on the selected process design factor. For this example:

$$J_{N\,des} = J_{N\,max}/DF \qquad (6\text{-}31)$$

$$J_{N\,des} = 2.0/1.5 = 1.33 \text{ g } NH_4^+\text{-N}/m^2/d$$

Other design parameters already established in this example are:
 $a = 138 \ m^2/m^3$
 $v_h = 80$ m/d
 $N_i = 21.0$ mg/L at r = 0
 $KN = 1.5$ mg/L

5. Determine the required media depth. Since k = 0, Equation 6-29 is applicable. To provide an additional margin of safety, design for an effluent ammonia-nitrogen concentration of 1.5 mg/L. For this example:

$$z = \frac{v_h}{a\ J_{N\,des}}\ [N_i - N + K_N \ln\left(\frac{N_i}{N}\right)] \qquad (6\text{-}32)$$

$$z = \frac{80}{(138)(1.33)}\ [21 - 1.5 + 1.5 \ln\left(\frac{21}{1.5}\right)] = 10.2 \text{ m (33.5 ft)}$$

6. Calculate the total biotower cross-sectional area required based on the desired hydraulic loading rate. The area is:

$$\text{Area} = \frac{Q}{v_h} = \frac{21,055 \ m^3/d}{80 \ m/d} = 263 \ m^2 \ (2,830 \text{ sq ft})$$

7. Determine the number of biotowers and their dimensions. Assume four equally sized biotowers are selected providing two parallel systems, which will each operate with two trickling filters in series. Assume that provisions will be made to reverse periodically the biotowers' order of operation in each of the two parallel systems. For this example, each biotower's area is:

Area = 263/2 = 131.5 m^2 (1,415 sq ft)

For circular systems the diameter of each biotower is:

$$\frac{\pi\ d^2}{4} = 131.5 \ m^2$$

$$d = 12.9 \text{ m (42.5 ft)}$$

Since each parallel train consists of two biotowers in series, the depth of each biotower is:

Depth = 10.2 m/2 = 5.1 m (16.7 ft)

8. Determine the media volume required. Media volume can be calculated from:

Media volume = (263)(10.2) = 4(131.5)(5.1) = 2,683 m^3 (94,740 cu ft)

6.5.2.3 Design Example No. 3 (continued)

9. Determine oxygen and air flow requirements. As discussed in Section 6.5.2.1, the recommended air flow requirements for combined BOD_5 removal and nitrification systems are given by (49,95):

$$m^3/min = 150(1.25 \, BOD_5, \, kg/d + 4.6 \, TKN, \, kg/d)(PF)/1,440$$

For design of tertiary nitrifying trickling filters receiving low levels of influent BOD_5, the recommendation was given that the minimum air flow should correspond to 50 kg O_2 supplied/kg O_2 consumed (49). Fifty kg of O_2 corresponds to 180 m^3 of air, which implies that this recommendation also was intended to include any additional COD removal that would occur when treating a high quality influent. The recommendation was also given that during peak loading periods, the oxygen supply should be 20 to 30 kg O_2/kg of O_2 used (49).

At average conditions, the nitrogenous oxygen demand (ignoring the small residual NH_4^+-N concentration) would result in an air flow requirement as follows:

$$\frac{(180)(21,055)(21)(4.6)}{(10^3)(1,440)} = 254.2 \, m^3/min \, (8,980 \, scfm)$$

The peak day nitrogen pollutant mass specified in Chapter 2 (Table 2-12) is 1.7 times the average day value. Even if there was no attenuation in the peak day loading by virtue of the preceding activated sludge system, an air supply of 254.2 m^3/min (8,980 scfm) would still supply 29.4 kg (64.8 lb) O_2/kg nitrogenous oxygen demand, which is within the range recommended for the peak loading period. Therefore, base the design of the aeration system on average conditions.

6.5.3 Rotating Biological Contactors

Rotating biological contactors (RBCs) have been used for treatment of municipal wastewaters since the 1960s. Most of the current installations are for carbonaceous BOD removal only. To a lesser extent, systems have been constructed for combined organic removal and nitrification or for separate stage nitrification.

Two previous EPA publications, *Design Information on Rotating Biological Contactors* (105) and *Review of Current RBC Performance and Design Procedures* (106), discuss the design and operation of RBCs in detail. The former publication presents a detailed design procedure for carbonaceous as well as nitrification applications. Little has changed in the design and construction of RBC facilities in the last 10 years. The design approaches described in these two publications are still commonly used and valid today. This section presents design procedures that are based on those discussed in these publications along with supplemental information from recent literature and from the major manufacturers.

6.5.3.1 Process Description

An RBC is a fixed film process in which plastic media are mounted as discs or spiral wound sheets to a rotating horizontal shaft. The shaft is mounted in a tank such that approximately 40 percent of the media surface is submerged during its rotation. The contactors are typically up to 3.6 m (12 ft) in diameter, with shaft lengths up to 8.2 m (27 ft). Size restrictions of the units are due primarily to transportation limitations.

The support medium has a surface configuration, generally unique to each manufacturer, that maximizes surface area and enhances turbulence and flow patterns across the surface. The media are generally available in two types: a standard medium provides approximately 9,290 to 9,660 m^2 (100,000 to 104,000 sq ft) of surface area per single shaft (3.6 m diameter by 7.6 m [12 ft by 25 ft] shaft length), and a high density medium provides surface areas of 13,940 to 14,490 m^2 (150,000 to 156,000 sq ft) for a similarly sized unit. Because spacing between medium surfaces in high density media is decreased, applications of these media are limited to situations where biomass is expected to be thin, such as in separate-stage nitrification systems or the last stages of carbonaceous or combined carbon oxidation-nitrification systems.

An RBC process train generally consists of several RBCs in series. The size limitation of an individual RBC unit requires two or more parallel trains to be employed for plants greater than 44 L/s (1 mgd) in capacity. Figure 6-23 presents a typical application of RBCs for municipal wastewater treatment.

The mass transfer of substrate (BOD_5 and/or ammonia-nitrogen) and oxygen is due to the rotation of the partially submerged discs through the wastewater contained in the
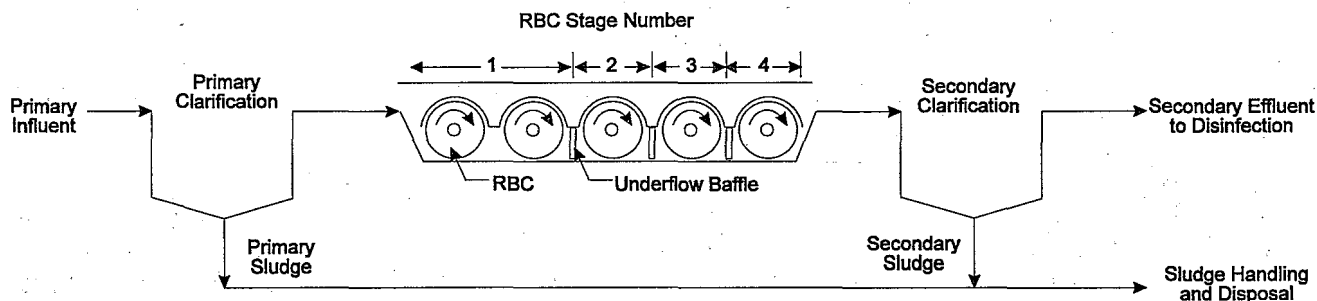
182

Figure 6-23. Typical application of rotating biological contactors to municipal wastewater treatment.

tank. The constant rotation of the media through the wastewater alternately exposes the biofilm, which develops on the media, to substrate and air. The shafts of full-scale RBCs (3.6 m [12 ft] diameter) are rotated at one to two revolutions per minute (rpm); a peripheral velocity of 18.3 m/min (60 ft/min) is most common (1.6 rpm). In addition to promoting substrate and oxygen diffusion, the rotation also regulates biofilm growth by sloughing excessive growth from the discs.

Excessive biofilm growth, generally the most common symptom of inadequately designed RBC systems, is attributable to excessive first-stage or total-system organic loading. Excessive growth reduces treatment efficiency by reducing biofilm surface area and limiting oxygen transfer. Low DO levels in the biofilm and bulk liquid promote the development of nuisance organisms, primarily *Beggitoia*, a sulfur-reducing bacteria. Such organisms can coat the biofilm surface, severely reducing the effective surface area available for carbon oxidation and nitrification. Excessive growth can also structurally damage shafts and media because of the added weight. The problems of excessive biological growth due to high organic loading and corresponding low DO in the first stage of RBC systems may be overcome by using a step-feed hydraulic flow pattern, an enlarged first stage, and/or supplemental aeration (107). Nitrification systems require low organic loads to promote the development of nitrifying bacteria. This requirement substantially reduces the potential for excessive growth development and prevents some of the common problems encountered by carbon oxidation systems.

All carbonaceous oxidation or combined carbon oxidation-nitrification RBC systems require secondary clarification for the removal of sloughed biofilm. While a recycle flow, such as is employed in activated sludge processes, is not necessary, recirculation may enhance nitrification performance by providing more optimum C:N throughout the system (108).

A sufficient biofilm normally develops in approximately two weeks from startup during normal operation at temperatures above 15°C (59°F). A normal biofilm for a combined carbon oxidation and nitrification system is generally a brown-gray film in the first and second stages and a thin, reddish-brown slime in the latter stages (109). The latter stages may also have a slight algal growth. The color of biofilm on separate stage nitrification systems has been characterized as tan to bronze (110). The color darkens as film thickness increases. Biofilm thickness reaches equilibrium in 25 to 60 days. In combined carbon oxidation-nitrification systems, thicker films are observed generally in the first and second stages where organic loadings are at a maximum and carbon oxidation occurs. Thinner films generally develop on the latter stages where nitrification occurs.

The active portion of the biofilm is controlled by the diffusion of oxygen and substrate into the film. The depth of active biomass has been reported to range from 20 to 600 μm (105); greater depths do not contribute to substrate removal. The development of excessively thick biofilms reduces active RBC surface area and affects treatment capacity. Biofilm thickness should be controlled to minimize excessive buildup.

Biofilm thickness can be indirectly monitored using load cells to weigh RBC shafts. These devices can be either permanently installed on the shaft or used periodically. Excessive biofilm is controlled by natural sloughing. Manual control is needed occasionally and is accomplished by reverse rotation, air injection, and/or increased rotational speed while the train is still in service. Out-of-service methods include caustic cleaning or chlorination.

6.5.3.2 Design Considerations

The two critical design parameters for RBC systems are the determination of disc area and staging requirements. While each manufacturer provides his own design curves and guidance, all are based on similar procedures. The basic information needed covers design loads (flow, organic, and nitrogen), load variability, and effluent limits.

6.5.3.3 Kinetics of Organic Substrate Removal

The organic substrate removal rates of many biological treatment processes for municipal wastewater are generally described as following first-order kinetics. The principal influence on the reaction rate is the complex nature of municipal wastewater. During exposure to the mixed culture that develops in most biological treatment systems, many compounds degrade readily, some degrade slowly, and others resist treatment.

Organic removal rates for RBCs have been characterized in the literature by zero-, first-, and second-order reactions. All manufacturers recommend staging RBCs when low effluent BOD_5 is required, which implies reaction kinetics greater than zero-order. Zero-order kinetics apply only when RBCs are meant to provide a low level or a roughing level of treatment (40- to 60-percent BOD_5 removal). In this case, although the removal of BOD_5 may follow a zero-order reaction rate, it is generally accepted that the availability of DO controls BOD_5 removal.

A first-order kinetic expression was developed based on an analysis of influent and RBC effluent data from five pilot and full-scale facilities (111,112). The expression is as follows:

$$\frac{S_e}{S_i} = e^{-K(V/Q)^{0.5}} \qquad (6\text{-}33)$$

where:

S_e = secondary effluent total BOD_5, mg/L
S_i = RBC influent total BOD_5, mg/L
V = media volume, m^3
Q = hydraulic loading, m^3/min
K = reaction constant

The media volume term (π by $radius^2$ by length) does not include the impact of total available media surface area, which was found to be a factor also affecting performance. This expression can be used to develop a family of design curves similar to those shown in Figure 6-24. Generally, first-order kinetics best describe design curves utilized by RBC manufacturers.

A second-order kinetic approach for predicting BOD_5 removal has been developed (105,113,114). This approach implies that the removal of soluble BOD_5 across a series of RBCs is proportional to the square of the residual soluble BOD_5 concentration. The equation is as follows:

$$C_n = \frac{1 + \sqrt{1 + 4kt(C_{n-1})}}{2\,kt} \qquad (6\text{-}34)$$

C_n = soluble BOD_5 in nth stage, mg/L
k = second-order reaction rate constant, L/mg/hr
t = hydraulic residence time in the nth stage, hr
C_{n-1} = soluble BOD_5 entering the nth stage, mg/L

The second-order reaction rate, k, was found to be 0.083 L/mg/hr based on an analysis of full-scale RBC data for municipal wastewater. This approach uses soluble BOD_5
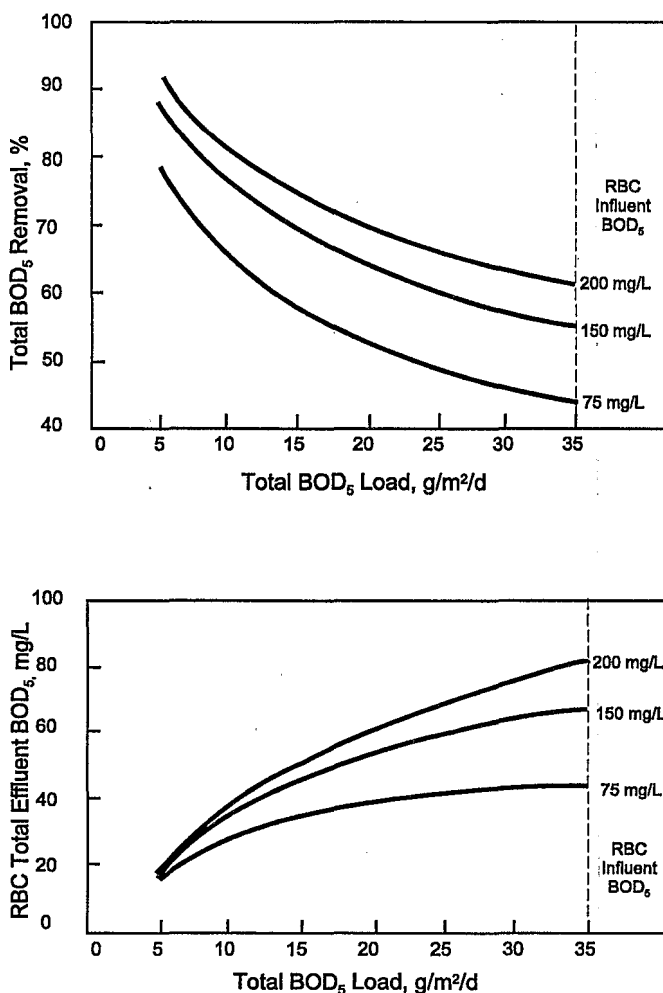

Figure 6-24. RBC design curves developed from Equation 6-33 (adapted from Reference 111).

for design and is not sensitive to primary clarifier performance relative to the level of solids (particulate BOD_5) entering the RBC system. Some consideration should be given to particulate BOD_5, which can exert an oxygen demand through entrapment of SS in the biofilm and their subsequent hydrolysis. The soluble BOD_5 within the system will thereby increase. A hydrolysis factor could be used, if known; alternatively, a maximum total BOD_5 loading condition can be imposed, which typically would be 30 g BOD_5/m^2/d (6 lb/1,000 sq ft/d). This second-order expression also can be used to develop a family of curves similar to those developed with the first-order approach.

Pilot studies, using full-scale equipment (3.6 [12 ft] m diameter disks), should be performed for wastewaters not considered "typical" municipal wastewaters. These would include industrial wastewater and municipal wastewater

with a significant industrial component. The results of the pilot studies can be analyzed, using the first- or second-order expressions, to develop a series of wastewater-specific design curves.

Design criteria developed through pilot studies using reduced-scale equipment should be used cautiously. Higher substrate removal rates have been reported for pilot systems with RBC diameters less than 3.6 m (12 ft). The higher rates are due to enhanced oxygen transfer efficiencies, which result from the higher rotational speeds needed to maintain peripheral velocities of 18.3 m/min (60 ft/min). Appropriate scale-up factors may be needed when pilot studies are used to develop design data.

6.5.3.4 Kinetics of Nitrification

Nitrification kinetics for RBCs are influenced by temperature, organic load, and effluent ammonia-nitrogen requirements. At NH_4^+-N concentrations greater than about 5 mg/L, the rate of nitrification for RBCs is oxygen-transfer limited down to a temperature of 13°C (55°F) (49). Below this temperature, corrections must be made to the nitrification rate. Most manufacturers provide design curves for temperature correction. Figure 6-25 is an example of these corrections for both nitrification and BOD_5 removal.

The development and balance of heterotrophic and autotrophic (nitrifying) organisms differs for combined and separate stage systems. In nitrifying activated sludge systems the nitrifier fraction of the mixed bacterial culture is controlled by the ratio of ammonia nitrogen to BOD. Similarly, the nitrifier fraction of the biofilm on an RBC is a function of the same ratio. This ratio will change with each preceding stage in an RBC system, which explains the

gradation in biofilm characteristics across an RBC train, especially in combined systems.

Organic load influences the rate of nitrification through the competitive development of heterotrophs and autotrophs. Autotrophs, which are slower growing than heterotrophs, can dominate the mixed culture population on a RBC only when the development of heterotrophs is substrate limited. The following relationship between organic loading and the rate of nitrification has been developed (116):

$$f_i = 1.43 - 0.1M \qquad (6\text{-}35)$$

when:

$$(4.3 < M < 14.3)$$

where:
f_i = fraction of maximum ammonia removal rate
M = organic substrate load, g $COD/m^2/d$

This relationship has been used to develop Figure 6-26. As shown, this analysis suggests that the nitrification rate approaches the maximum rate when the BOD_5 loading is less than approximately 3.5 g $BOD_5/m^2/d$ (0.7 lb/1,000 sq ft/d).

A significant degree of nitrification will not occur on RBCs until the soluble BOD_5 is less than 15 mg/L. This phenomenon is due to the autotroph/heterotroph population dynamics. A review of Figure 6-24, with extrapolation of the effluent BOD_5 vs. loading curves to zero, shows that 15 mg/L BOD_5 is reached at loadings of approximately 3.5 g/m²/d (0.7 lb/1,000 sq ft/d). This agrees with the previous analysis of Equation 6-35 (Figure 6-26).
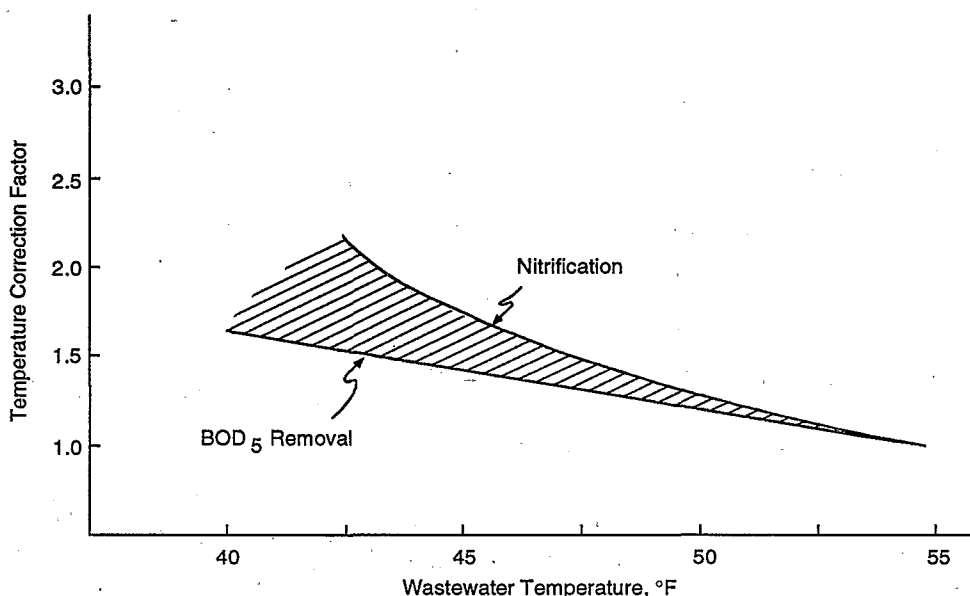

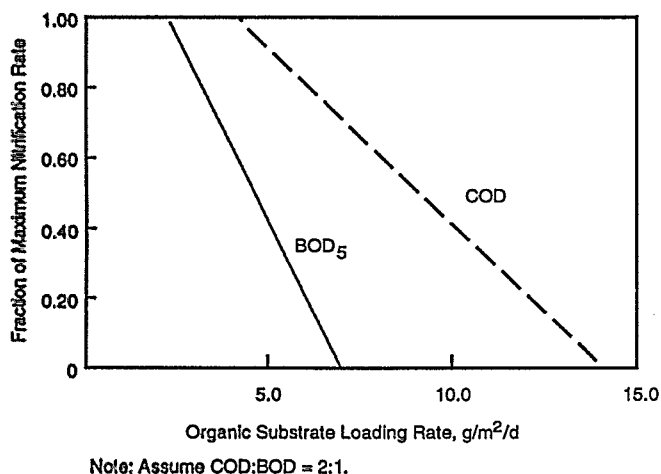Figure 6-25. Temperature correction factors for required RBC area (adapted from Reference 115).

185

Figure 6-26. Effect of organic substrate loading on the rate of nitrification (adapted from Reference 116).

The rate of nitrification has also been shown to approach a first-order relationship with respect to the effluent concentration at effluent ammonium-nitrogen levels less than 5 mg/L; the maximum zero-order removal rate occurs at levels above 5 mg/L. This function is demonstrated graphically in Figure 6-27, which uses data from three municipal nitrifying RBC plants (105). Manufacturers have incorporated this into their design procedures by specifying nitrogen removal rates as a function of effluent requirements.
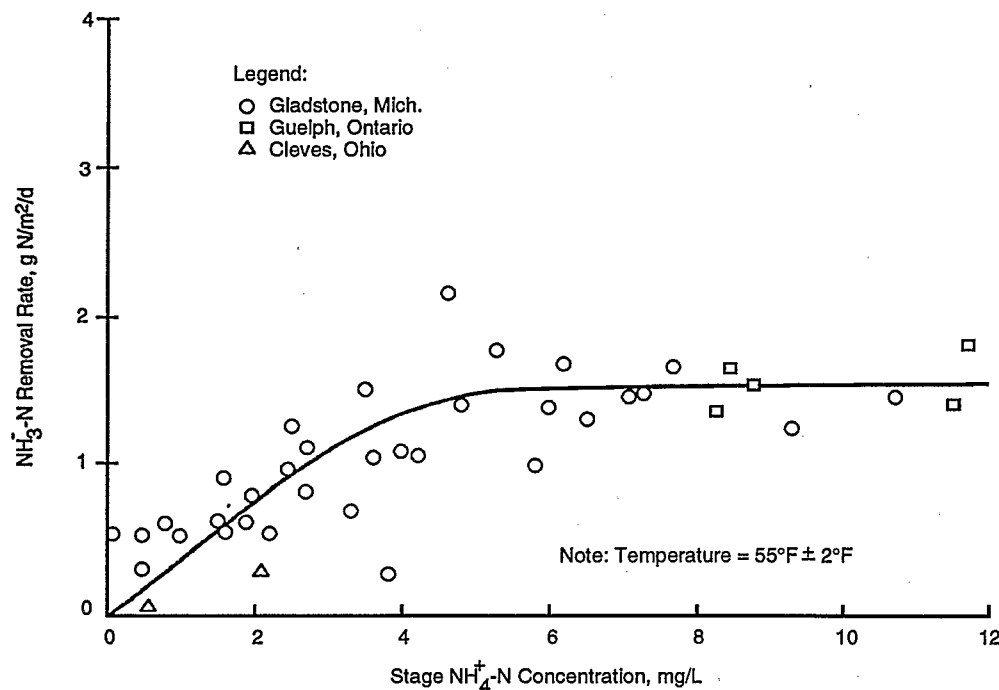
6.5.3.5 Empirical Procedures

Empirical design approaches are still the most common design procedures used for determining RBC media area requirements. Fundamental design procedures and mathematical models have been developed that incorporate kinetic parameters, mass transfer of substrate and DO, and changes in RBC physical configuration due to biofilm growth and sloughing. These fundamental approaches are complex and often have not been adequately demonstrated; they are generally not used for design purposes because of their complexity.

The prevalent empirical design approaches are based on substrate and hydraulic areal loading rates. Substrate loading rates control oxygen demand and biofilm thickness. Substrate loading parameters for carbon oxidation systems include total and soluble BOD or COD, while loading parameters for nitrification systems include ammonium nitrogen, TKN, and soluble organic nitrogen. Generally, nitrification designs should use TKN as the basis for sizing RBCs. Hydraulic loading rates are generally based on specific influent substrate concentrations; as such, they indirectly become a substrate loading rate.

First-stage organic loading conditions are considered the most critical factor in the design and operation of RBCs for carbonaceous removal or combined carbon oxidation-nitrification. Substrate loadings and oxygen demand are at maximum levels in the first stage. Oxygen demand due to excessive substrate loadings can easily exceed oxygen

Figure 6-27. Full-scale RBC nitrification rates at a wastewater temperature of 13°C (55°F) (adapted from Reference 105).

186

transfer capacity if the first stage is not sized properly. It is not unusual for the first stage of a properly designed carbonaceous oxidation system to contain up to 50 percent of the total system's surface area.

The principal design criterion, with respect to organic removal, is not to exceed soluble BOD_5 loadings of about 12.2 $g/m^2/d$ (2.5 lb/1,000 sq ft/d) and total BOD_5 loads of 31.2 $g/m^2/d$ (6.4 lb/1,000 sq ft/d) for any individual stage. Maximum oxygen transfer rates for full-scale RBCs were reported to be 6.8 to 7.3 g $O_2/m^2/d$ (1.4 to 1.5 lb/1,000 sq ft/d) (117). These rates correspond to a maximum BOD removal capacity of 13.7 to 14.6 $g/m^2/d$ (2.8 to 3.0 lb/1,000 sq ft/d) at a presumed COD:BOD of 2:1. Thus, the maximum soluble BOD_5 removal rate generally referred to in the literature, 12.2 $g/m^2/d$ (2.5 lb/1,000 sq ft/d), is slightly less than the rate that can be supported by oxygen transfer.

The maximum oxygen transfer rate of 6.8 to 7.3 g $O_2/m^2/d$ (1.4 to 1.5 lb/1,000 sq ft/d) also translates to a maximum nitrogen removal rate of approximately 1.45 to 1.61 g $N/m^2/d$ (0.3 to 0.33 lb/1,000 sq ft/d), which agrees with the maximum zero-order rate shown in Figure 6-27. The empirical design approaches used by many manufacturers are based on maximum substrate loading rates and design curves similar to those in Figure 6-24.

6.5.3.6 Staging Requirements

Substrate (organic and nitrogen) removal by RBCs is generally described as a zero- or higher-order reaction. Staging becomes important when low effluent BOD_5 or ammonia-nitrogen concentrations are required; a point at which the removal rate is often described as a first-order reaction. Many of the manufacturers recommendations for staging are a function of substrate removal or effluent concentration (115,118–120). Table 6-24 presents examples of manufacturer recommendations.

6.5.3.7 Load Variability Considerations

RBC systems are typically designed for average flow and loads (BOD_5 and ammonia nitrogen) when the ratio of peak to average design flows is less than 2.5 to 1. In cases where the ratio is greater than 2.5 to 1, the average flow/substrate design loading should be reduced by a factor proportional to the increased variability above the 2.5:1 ratio. Alternatively, flow equalization should be incorporated into the process train.

6.5.3.8 Flow Distribution

Flow distribution is an important design consideration for RBC systems that employ several parallel trains. The lack of positive flow measurement and control promotes maldistribution of flow among the trains, leading to overloaded and underloaded RBC trains within the same system. Poor flow distribution results in variable biofilm thickness among the first and middle stages of parallel systems.

6.5.3.9 Design of Combined Systems

The most common application of RBCs for nitrification is for combined carbon oxidation and nitrification, in which the media area requirement is the sum of the media requirements for carbon oxidation and nitrification. The design of these systems must consider organic loading as well as nitrogen loading. Insufficient first-stage RBC area leads to the development of nuisance organisms, which can effectively reduce the active media area. When this occurs in the first stage, additional organic removal must then be accomplished in the latter stages. Nitrification activity is displaced further down the train. If there is insufficient stage/area capacity, nitrification performance may fail. As such, proper sizing of the RBC system for carbon oxidation is critical for successful nitrification. Sizing should be checked against maximum expected loads to ensure that recommended stage and total system loading are not exceeded under these conditions.

Table 6-24. Manufacturer Recommendations for RBC Staging (Adapted from Reference 105)

		Carbon Oxidation		Nitrification	
		Effluent BOD_5	Number Stages	Effluent Ammonia-N	Number Stages
Envirex (118)		>25 mg/L	1	5 mg/L	1
		15–25 mg/L	1–2	<5 mg/L	Based on first-order kinetics
		10–15 mg/L	2–3		
		<10 mg/L	3 or 4		
Lyco (115)		<40% removal	1	<40% removal	1
		35–65% removal	2	35–65% removal	3
		60–85% removal	3	60–85% removal	3
		80–90% removal	4	80–95% removal	4

One design approach described in this section parallels the approach specified by a major manufacturer. Other manufacturers' design methods may differ slightly, but most depend on using typical areal loading rates (hydraulic and substrate mass) and design curves tempered with maximum loading rates.

When designing a combined system, the sizing of the first stage for carbon oxidation is critical. The first stage should be designed for a maximum total BOD_5 loading of 31.2 $g/m^2/d$ (6.4 lb/1,000 sq ft/d). The soluble BOD_5 loading should not exceed 12.2 to 19.5 $g/m^2/d$ (2.5 to 4.0 lb/1,000 sq ft/d). Under these loading conditions, a BOD_5 reduction of 40 to 50 percent can be expected, based on a maximum BOD_5 removal rate of 13.7 to 14.6 $g/m^2/d$ (2.8 to 3.0 lb/1,000 sq ft/d). Conservative design would limit the soluble BOD_5 loading to 12.2 $g/m^2/d$ (2.5 lb/sq ft/d). ,

The total area for subsequent carbonaceous oxidation stages is then determined based on the first-stage effluent BOD_5 load and a target soluble BOD_5 concentration of 15 mg/L from the last stages of the carbon oxidation portion of the total system. In other words, one wants a soluble BOD_5 of less than 15 mg/L before entering the nitrification stages. This removal efficiency is used with the first stage effluent BOD_5 concentration to determine the design hydraulic loading ($m^3/m^2/d$) from design curves

similar to Figure 6-28. The wastewater flow and the hydraulic loading rates are then used to determine the RBC area required for carbonaceous removal for the subsequent stages. The total area requirement for carbon oxidation is the sum of the first stage and subsequent stage areas. Some manufacturers provide design curves based on effluent BOD_5 concentration rather than on percent removed. These curves should be used in the same manner to determine design loadings.

RBC area requirements for nitrification should be based on TKN loads. Nutrient uptake and effluent loads should be subtracted from the influent nitrogen load to determine the mass of nitrogen to be removed. The overall nitrification rate is selected based on the target effluent ammonium-nitrogen concentration. One approach to computing the additional media area required for nitrification is use of a manufacturer supplied design curve such as Figure 6-29. Alternatively, the relationship of the nitrification rate to effluent ammonium nitrogen should be similar to Figure 6-27; the rate data in this figure and the mass of nitrogen to be removed provide another method to compute the RBC area required for nitrification.

The total area requirement is the sum of the areas required for carbon oxidation and nitrification. For combined carbon oxidation/nitrification applications, a minimum of
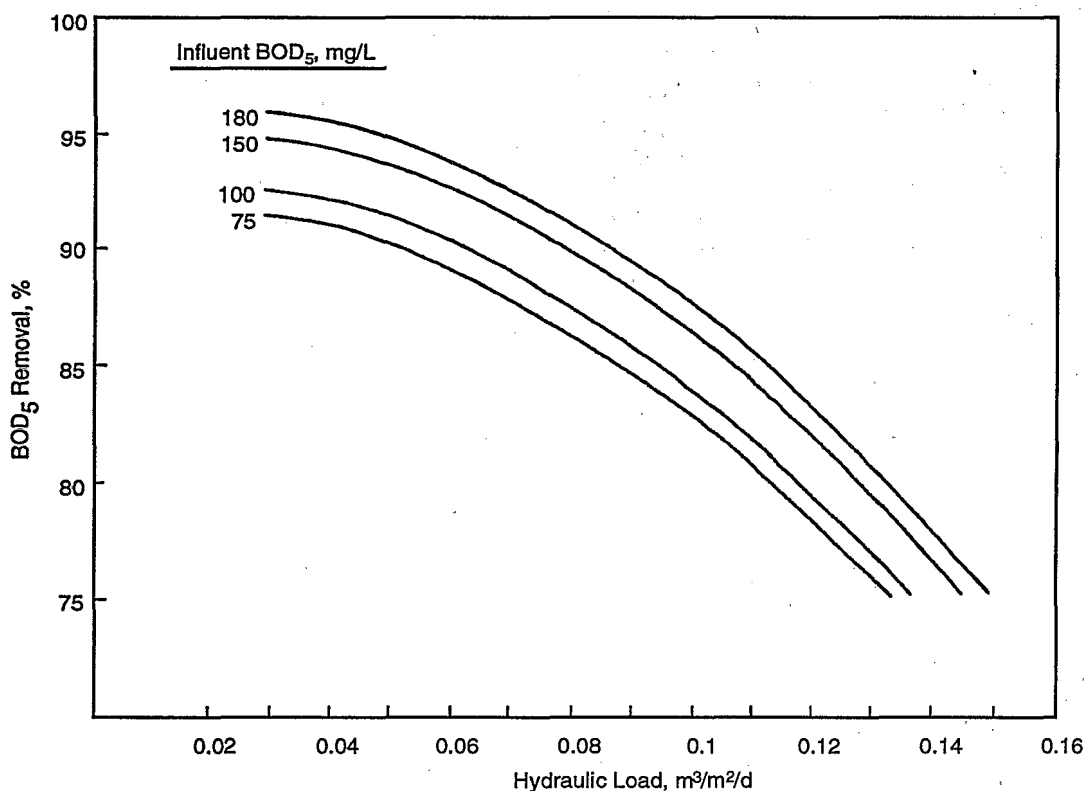

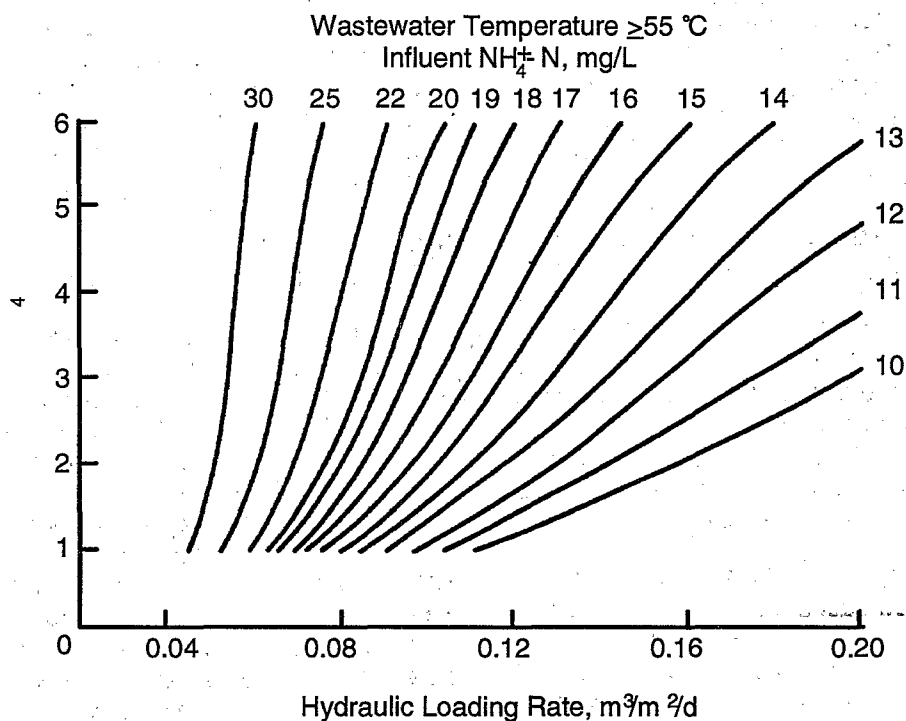Figure 6-28. Typical RBC design curve (adapted from Reference 118).

188

Wastewater Temperature ≥55 °C
Influent NH$_4^+$-N, mg/L

Hydraulic Loading Rate, m³/m²/d

Figure 6-29. Nitrification design relationships (adapted from Reference 121).

four stages is generally required. Interstage loads should be considered when RBC disc area is being allocated to the stages. Interstage loads beyond the first stage should not exceed 29 g TBOD$_5$/m²/d (6 lb TBOD$_5$/1,000 sq ft/d) or 12.2 g SBOD$_5$/m²/d (2.5 lb SBOD$_5$/1,000 sq ft/d).

Area and staging requirements should be developed for summer and winter conditions, and should consider seasonal effluent limits. RBC area requirements for carbon oxidation and nitrification should be temperature corrected separately, based on manufacturer supplied procedures such as shown in Figure 6-25. The design should be based on the condition requiring the most surface area.

6.5.3.10 Design Approach for Separate-Stage Nitrification

The design of RBCs for separate-stage nitrification is similar to the nitrification design for combined carbon oxidation-nitrification systems. Influent soluble BOD$_5$ to a separate-stage process should be less than 15 mg/L; otherwise, additional surface area should be provided for carbon oxidation. Design should be developed for winter and summer conditions, with the condition requiring the most area controlling final design. Provision for periodic flow reversal (e.g., once a week) can also lead to a higher capacity to handle diurnal NH$_4^+$-N fluctuations because of higher nitrification potential in the last and usually NH$_4^+$-N limited stage(s) (122).

6.5.3.11 Design Example No. 4: Nitrification with an RBC System at High Carbonaceous Feed Concentration

The following design example illustrates RBC design for a combined carbon oxidation-nitrification system. Effluent limits are for the more stringent criteria specified in Table 2-10. Assume that the plant contains primary clarification, anaerobic sludge digestion and tertiary filtration such that the primary effluent characteristics and overall plant flow are the same as summarized in Table 2-16. A summary of wastewater characteristics, effluent limits, and design conditions and assumptions is presented in Table 6-25.

Table 6-25. Design Conditions for Example No. 4

Wastewater Flow Characteristics, m^3/d (mgd)		
Raw wastewater average flow	18,925 (5.0)	
Total influent average flow	21,055 (5.5)	
Primary Effluent Characteristics, mg/L		
COD	187[a]	168[b]
$CBOD_5$	97[a]	87[b]
Soluble $CBOD_5$	53[a]	48[b]
TSS	80[a]	72[b]
TKN	29.5[a]	26.6[b]
Alkalinity, mg/L (as $CaCO_3$)		120[b]
Secondary Effluent Permit Limits, mg/L		
CBOD5	10[c]	
TSS	10[c]	
NH_4^+-N	2[c]	
Total N	5[c]	
Design Conditions/Assumptions		
Reactor temperature, °C	15	
Reactor pH range	7.0–7.6	
Nitrogen available for nitrification, mg/L equivalents	24.5 (22.0 mg/L)[b]	

[a] Concentration value at average conditions expressed as mg/L equivalents (Table 2-16).
[b] Concentration value at average conditions at total flow of 21,055 m^3/d.
[c] From Table 2-10 for year-round monthly maximum permit limits.

1. *Evaluate peaking factor requirement.* Since the peak to average flow ratio is less than 2.5:1, the design loadings will be based on average conditions.

2. *Evaluate temperature correction.* Since the design temperature is greater than 13°C (55°F), the temperature correction factor from Figure 6-25 for both carbonaceous oxidation and nitrification is 1.0.

3. *Determine media requirements to reduce soluble BOD_5 to 15 mg/L.* Do not exceed loading limits in the first stage of 29 g total BOD_5/m^2 (6 lb/1,000 sq ft) or 12.2 g soluble BOD_5/m^2 (2.5 lb/1,000 sq ft).

(18,925)(97)/29 = 63,300 m^2 (681,400 sq ft) required based on total BOD_5

(18,925)(53)/12.2 = 82,215 m^2 (885,000 sq ft) required based on soluble BOD_5

A minimum of nine standard density shafts (9,290 m^2 [100,000 sq ft]) are required for the first stage providing a total of 83,610 m^2 (900,000 sq ft) of surface area. This minimum requirement corresponds to a first-stage loading of:

$$\frac{21,055 \ m^3/d}{(9)(9,290 \ m^2)} = 0.25 \ m^3/m^2/d$$

When using figures such as Figure 6-28 for design, a common assumption is that effluent BOD would be 50-percent soluble. Hence to achieve a soluble BOD_5 of 15 mg/L would require an overall BOD_5 removal

6.5.3.11 Design Example No. 4 (continued)

of 65 percent. Based on an extrapolation of the influent BOD_5 curves shown in Figure 6-28, a hydraulic load of about 0.15 m/d should be adequate. Therefore, a total of 140,400 m² (1,511,000 sq ft) of media surface is required.

Initally assume a total of five trains with two shafts per train in each of the first stages and one shaft per train in the second stage. Check the design against the predictions from Equation 6-34 to see if there is reasonable agreement. At the standard 0.0049 m³/m² of media (0.12 gal/sq ft) tank sizing, the hydraulic residence time in the first and second stages is as follows:

$$t_1 = \frac{(0.0049)(9,290 \text{ m}^2/\text{shaft})(2)(24)}{21,055/5} = 0.52 \text{ hr}$$

$$t_2 = \frac{t_1}{2} = 0.26 \text{ hr}$$

From Equation 6-34:

$$C_1 = \frac{-1 + \sqrt{1 + 4(0.83)(0.052)(48)}}{2(0.52)(0.083)} = 23.7 \text{ mg/L}$$

$$C_2 = \frac{-1 + \sqrt{1 + 4(0.083)(.26)(23.7)}}{2(.26)(0.83)} = 17.3 \text{ mg/L}$$

Consider modifying the design to a total of 18 shafts and 6 trains to ensure adequate carbonaceous oxidation. Under these conditions, Equation 6-34 would predict soluble BOD_5 concentrations of 22.3 and 15.8 mg/L from Stages 1 (12 shafts) and 2, respectively. Additional safety would be provided to deal with the sustained peak month loading.

4. *Determine additional media requirements for nitrification.* The soluble, nondegradable TKN from the RBC equals about 1.0 mg/L and the TKN associated with the waste biological solids equals approximately 3.0 mg/L. The permit limit is 2 mg/L of NH_4^+-N; to provide a margin of safety this design is for an effluent concentration of 1 mg/L. Assume that the equivalent NH_4^+-N concentration available for nitrification is equal to 22.6 mg/L (26.6 − 4) at the total influent flow including plant recycles of 21,055 m³/d.

For an effluent NH_4^+-N concentration of 1 mg/L, a hydraulic loading of about 0.057 m³/m²/d (1.4 gpd/sq ft) would be adequate, according to Figure 6-29. An additional media requirement of 369,400 m² (3.98 x 10⁶ sq ft) results, which corresponds to an additional 40 shafts of standard density media (9,290 m² or 100,000 sq ft per shaft) or 27 shafts of high density media (13,940 m² or 150,000 sq ft per shaft).

As an alternate approach, Figure 6-27 indicates a maximum nitrification rate of 1.46 g NH_4^+-N/m²/d (0.3 lb/sq ft/d) would be observed down to about 5 mg/L of NH_4^+-N. Hence to reduce an influent concentration of 22.6 mg/L to 5 mg/L requires:

$$\frac{(22.6 - 5)(21,055)}{1.46} = 253,800 \text{ m}^2 \ (2.73 \times 10^6 \text{ sq ft})$$

At 1 mg/L NH_4^+-N, the reaction rate from Figure 6-27 is only about 0.5 g/m²/d (0.1 lb/1,000 sq ft/d). To reduce 5 mg/L NH_4^+-N to 1 mg/L at this reaction rate (i.e., the rate that would be observed in one complete mix stage) would require an additional:

$$\frac{(5 - 1)(21,055)}{0.49} = 171,880 \text{ m}^2 \ (1.85 \times 10^6 \text{ sq ft})$$

The total requirement for nitrification would be 425,700 m² (4.58 x 10⁶ sq ft). This translates to 46 standard density shafts or 31 high density shafts.

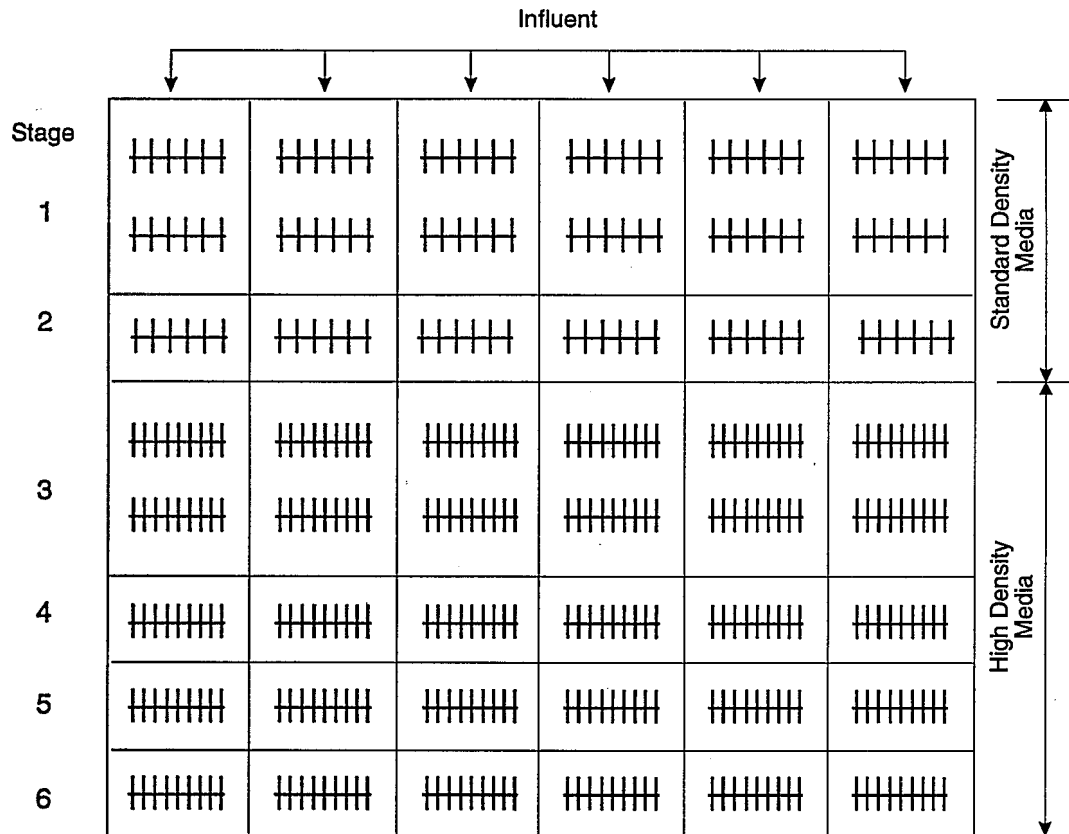# 6.5.3.11 Design Example No. 4 (continued)

Influent

Figure 6-30. RBC configuration for Design Example 4.

Figure 6-30 provides a convenient layout for this design example showing 18 shafts for carbonaceous BOD_5 removal down to a soluble BOD_5 of 15 mg/L and 30 high density shafts for NH_4^+-N removal. Combining two shafts in the first stage prevents organic overloading and placing two shafts in the third stage will not affect NH_4^+-N removal rates since the bulk liquid NH_4^+-N concentration is >5 mg/L.

It is interesting to reexamine the nitrification design shown in Figure 6-30 on a stage-by-stage basis. Assuming the maximum nitrification rate of 1.46 g NH_4^+-N/m^2/d (0.3 lb/1,000 sq ft/d) through Stages 3 and 4 (18 shafts total), the NH_4^+-N concentration would be reduced from 22.6 mg/L to 5.2 mg/L. From Figure 6-27, the concentration in Stage 5 corresponding to the following condition:

$$(\text{removal rate})(6)(13,940) = (5.2 - C_e)(21,055)$$

is 2.1 mg/L at a removal rate of 0.77 g NH_4^+-N/m^2/d. In Stage 6, the concentration corresponding to the following condition:

$$(\text{removal rate})(6)(13,940) = (2.1 - C_e)(21,055)$$

is 0.86 mg/L at a removal rate of 0.31 g NH_4^+-N/m^2/d. This type of stage-by-stage analysis of a preliminary design will produce a more accurate estimate of effluent NH_4^+-N concentrations than just assuming one rate at the desired effluent concentration as was done earlier in this design example. Whether or not this increased accuracy is of practical significance is an entirely different matter. This type of analysis demonstrates that fixed film RBC systems increase in area dramatically as effluent ammonia concentrations become very stringent.

5. *Check alkalinity requirements.* Nitrification of 21.6 mg/L of NH_4^+-N will consume 153 mg/L of alkalinity as $CaCO_3$. To maintain a residual alkalinity of 50 mg/L as $CaCO_3$ requires:

153 + 50 − 120 = 83 mg/L

From Table 2-3, the lime addition required is:

$$\frac{(83)(21,055)}{1.8(10^3)} = 970 \text{ kg } (2,140 \text{ lb) quicklime/d}$$

6. *Account for the effect of wastewater temperature.* When using the design curves presented in Figures 6-28 and 6-29 for temperatures less than 13°C (55°F), the required media surface areas need to be adjusted by the factors indicated in Figure 6-25; for temperatures above 13°C (55°F), the design curves do not need to be adjusted. Some data show that nitrification rates may increase to near 2 $g/m^2/d$ (0.4 lb/1,000 sq ft/d) at temperatures near 18.3°C (65°F ± 5°F), but there was enough variation in the data so that assuming higher rates would not be justified in the absence of pilot scale data (105).

7. *Reduce the nitrate nitrogen to an acceptable level.* To achieve the final effluent limit of 5 mg/L total nitrogen specified in Table 6-25 requires the addition of a separate-stage denitrification system to reduce the nitrate nitrogen to an acceptable level. Design of such systems is detailed in Chapter 7.

6.5.4 Other Attached Growth Reactors

Packed-bed reactors, biological aerated filters, and fluidized-bed reactors represent attached growth processes that have been utilized 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.

Information pertaining to use of these attached growth reactors for nitrification of municipal wastewaters is provided in Sections 6.5.4.1 through 6.5.4.3.

6.5.4.1 Packed-Bed Reactors

A packed-bed reactor, often referred to as a submerged filter, contains a stationary bed of media which 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 (123) or high-purity oxygen (124) 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 prior to the feed entering the reactor (123).

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 (1). 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 (123–127) 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.

In 1975, it was reported that 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 (123–127). 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:

- The use of media with a high void volume (greater than 90 percent).

- The supply of oxygen by the direct introduction of air into the bottom of the reactor.

- Operation at low influent $CBOD_5$ and SS.

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 (123–126), onsite piloting is recommended if the technology is being considered for use on a full scale.

6.5.4.2 Biological Aerated Filters

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. As noted in an EPA-sponsored study (128), 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 coarser, low density media is utilized.

- 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 (128). 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 back-

washing 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. The backwash water is normally 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 Figure 6-31. Biocarbone is the trademark name given to Omnium de Traitement et de Valorisations (OTVs) commercial embodiment of the process.

The primary advantage of the BAF is biotreatment 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.

The first commercial, full-scale BAF system began operation in 1982 in Soissons, France (129). Since that time a number of systems have been installed in Europe, Japan, and North America (128,130). As of 1990, there were approximately 30 commercial full-scale Biocarbone BAF systems installed or under construction, designed at wastewater flows of 1,900 22 L/s (0.5 mgd) or greater (130). The largest Biocarbone BAF system installed to date is designed to treat approximately 1,056 L/s (24 mgd) (129). 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.

The original media employed in the Biocarbone BAF was granular activated carbon. 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 (128). Biodamine is an angular shaped media whereas Biodagene is more spherical.
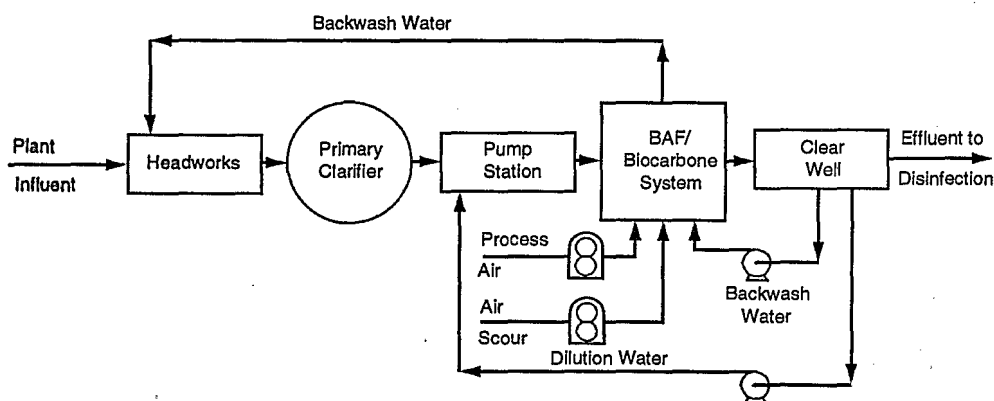

Figure 6-31. Biocarbone BAF system example flowsheet (adapted from Reference 128).

The angularity and size range of the media significantly affects the BAF treatment performance and operating requirements. The use of smaller media (i.e., 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 (128). The smaller media has been recommended when nitrification is required (130).

Process design and performance information on the general application of the BAF to municipal wastewater treatment can be found in other publications (128,131,132). Included is information on the questions of aeration and backwashing requirements, solids production, and design of system components. The following discussions present information on nitrification applications.

Pilot plant studies by the developer of the Biocarbone BAF system 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 (Figure 6-32) indicate that at a COD volumetric loading above approximately 3.2 $kg/m^3/d$ (200 lb/1,000 cu ft/d), nitrification is substantially reduced because of increased heterotrophic organism growth and associated oxygen consumption.

Data from a Biocarbone pilot plant study are presented in Figure 6-33 (130). The results depict the performance of a 140-L (37-gal) BAF pilot unit receiving primary municipal wastewater. A COD volumetric load of less than 2.0 $kg/m^3/d$ (125 lb/1,000 cu ft/d) was required to achieve approximately 90-percent 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). Reactor temperature information was not provided for the results presented in Figures 6-32 and 6-33.

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 (132), the BOD_5 volumetric loading must be limited to approximately 1 $kg/m^3/d$ (62.4 lb/1,000 cu ft/d) to achieve near 90-percent ammonium oxidation in a single unit. This conclusion is based on 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 effluent, the COD volumetric loading must be limited to approximately 5 $kg/m^3/d$ (312 lb/1,000 cu ft/d). The volumetric loading rate results indicate that carbonaceous oxidation and nearly complete nitrification of primary treated wastewater can
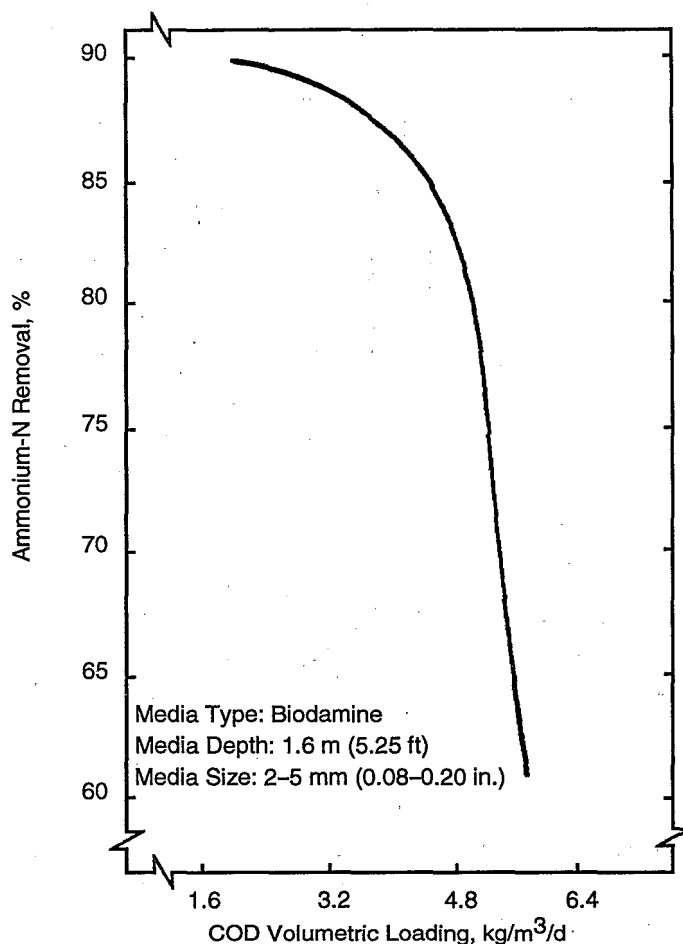

Figure 6-32. Effect of COD volumetric loading on Biocarbone BAF nitrification performance in the treatment of primary effluent (adapted from Reference 128).

be achieved in single BAF units at an empty-bed hydraulic retention time of approximately 1.5 to 3.5 hours.

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 (128), 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 time of 0.83 hr 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 $kg/m^3/d$ (36 lb/1,000 cu ft/d). Other reports indicate that over 90-percent removal of ammonium nitrogen is achievable at comparable volumetric loading rates at temperatures as low as 13.5°C (56.3°F) (129,133).
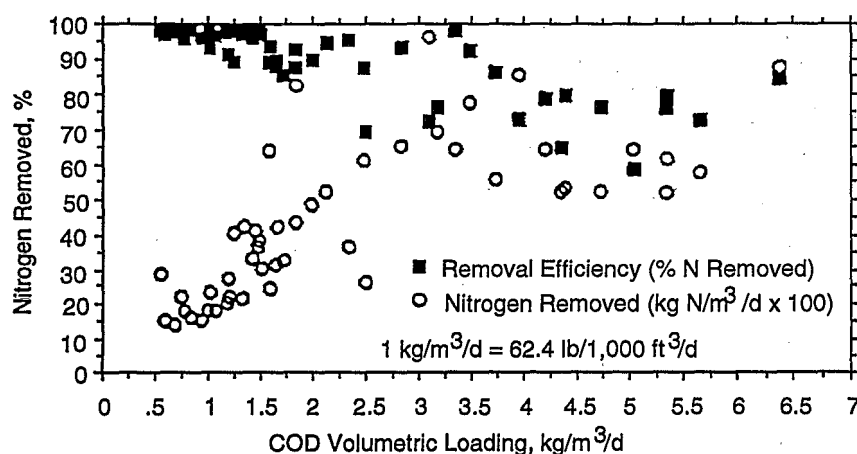
Figure 6-33. Effect of COD volumetric loading on nitrification of primary treated wastewater in a Biocarbone BAF pilot unit (adapted from Reference 130).

Although full-scale application of BAFs for municipal wastewater treatment has become widespread in recent years, particularly in Europe (130), the amount of operating and performance information on U.S. installations is limited. The lack of an extensive data base on nitrification applications suggests that onsite piloting may be warranted before selecting a technology.

6.5.4.3 Fluidized-Bed Reactors

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 5 to 10 times greater than that normally maintained in a conventional suspended growth reactor. 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:

- The device or method to distribute the influent flow to the reactor.

- 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).

- The device or method to control the expansion of the fluidized bed due to 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 design of the critical components have been presented elsewhere (134).

Although the development of water and wastewater systems using a fluidized bed of biomass can be traced back to the 1940s in England (135), 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 U.S. patent in 1974 (assigned to Ecolotrol, Inc.) for the application of the fluidized-bed process configuration to "denitrifying wastewater" (136). 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 (137).

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. 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 (138). Although hailed at that time as the most significant devel-

196

opment in the wastewater treatment field in the last 50 years, it also was claimed that no full-scale plants were yet in operation (139). 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 mechanical scale-up issues, slow development of economically attractive system configurations, and proprietary constraints (140). •

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 (140). Although full-scale fluidized-bed industrial systems are operating under conditions that result in nitrification, 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 (Section 7.4).

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 (141 to 147) is summarized as follows:

• A half-order model appears appropriate to describe the kinetics of ammonium oxidation in fluidized-bed reactors under nonlimiting DO conditions.

• The volumetric removal rate and the specific ammonium oxidation rate decrease significantly at low reactor ammonium concentrations.

• 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 minutes. This HRT is for treatment of municipal 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.

6.6 Combined or Coupled Suspended and Attached Growth Systems

6.6.1 General

A variety of approaches combine suspended and attached growth components into a single treatment process. For purposes of this section, a single treatment process means any system in which the wastewater contacts both a suspended growth component and an attached growth component prior to final settling. For example, a roughing filter may precede an activated sludge system or a trickling filter may precede a solids contact reactor. In each of these examples, an attached growth and suspended growth system have been combined. In other systems, called coupled systems, suspended and attached growth biomass operate within the same reactor(s). In coupled systems, settled solids from the final clarifier may or may not be recycled back through the fixed biomass component in each application, although this option is available in all the coupled systems to be discussed in this section.

Combining or coupling suspended and attached growth systems can offer several process advantages including protection against biomass washout, improved handling of industrial discharges or toxic shock loads, improved SVIs and SS settling velocities, and overall ease of operation. This section describes approaches used for nitrification, including:

• Trickling filter/activated sludge (TF/AS)

• Trickling filter/solids contact (TF/SC)

• Aerated biological filter/activated sludge (ABF/AS)

• Coupled Systems
  – Captor
  – RingLace
  – Linpor
  – FAST
  – Bio-2-Sludge
  – Monitor

Fundamental aspects of these systems are presented along with selected data about several of them.

6.6.2 Trickling Filter/Suspended Growth

Trickling filter/suspended growth processes encompassing a variety of applications have been summarized (148). According to some reports, the trickling filter may serve as a roughing filter with typical loadings of 1.6 to 3.2 kg $BOD_5/m^3/d$ (100–200 lb BOD/1,000 cu ft/d) placed immediately upstream of an activated sludge system. Or the trickling filter may be sized to ensure the necessary level of nitrification ahead of a solids contact reactor. In either

configuration, recycle solids from the final settler are returned to the suspended growth reactor only.

To distinguish the TF/SC process from TF/AS, EPA defined TF/SC as having the following distinguishing features:

- The primary function of the solids contact tank is to increase solids capture and particulate BOD removal.

- The majority of the soluble BOD removal occurs in the trickling filter.

- Return sludge solids are mixed with the trickling filter effluent.

- The solids contact tank is not designed to nitrify (the solids residence time is less than approximately two days) although nitrification may occur in solids contact if the trickling filter is the source of the nitrifying organisms.

- The aerated solids contact time is one hour or less based on total flow, including recycle.

In yet another process variation (ABF/AS), the settled solids are recycled through the trickling filter (Figure 6-34) to achieve both a coupled system within the filter as well as a combined trickling filter/activated sludge system.

6.6.2.1 Trickling Filter/Activated Sludge

The most common application of the TF/AS configuration, where nitrification must be achieved, is to use the trickling filter as a "roughing filter" for partial oxidation of the incoming carbonaceous material and to design the activated sludge system with a sufficient solids residence time to ensure that nitrification can be sustained. With this approach the roughing filter can be designed to achieve any level of treatment desired. Alternatively, for upgrading of existing trickling filter facilities, the benefits of existing infrastructure may be retained. This upgrading approach was taken at existing trickling filter plants in Livermore and Lompac, CA, Corvallis, OR, and El Lago,

TX (149). Factors such as odor production, ventilation, or media clogging will control maximum feasible trickling filter loadings. Irrespective of the initial trickling filter loading in the range of 0.74 to 3.2 kg $BOD_5/m^3/d$ (46 to 200 lb $BOD_5/1,000$ cu ft/d), the solids residence times in the activated sludge system controlled the final effluent quality with regard to soluble BOD and control of nonsettleable influent solids (150).

A survey of seven roughing filter/activated sludge facilities found a mean trickling filter design loading of 0.78 kg $BOD_5/m^3/d$ (48 lb $BOD_5/1,000$ cu ft/d) (148). Organic loadings can be much higher. Data from plastic media trickling filter studies at Sacramento, California, indicate oxidation in the tower was about 40 percent of the applied BOD_5 over the loading range of 1.6 to 6.4 kg $BOD_5/m^3/d$ (100 to 400 lb $BOD_5/1,000$ cu ft/d), with oxidation determined by the difference between influent and unsettled effluent BOD_5 values (151). Removals in the 1.3 m (4.25 ft) deep rock filter with 5 to 10 cm (2 to 4 in) media at Livermore, California, were not as good, even at the lower organic loadings shown in Table 6-26.

The unsettled wastewater exiting the trickling filter consists of a mixture of incoming soluble BOD not metabolized in the filter, incoming particulate degradable SS not precipitated or bioflocculated within the filter, active sloughed biomass and associated captured particulates not yet metabolized, and biomass decay products and other inert components in the entering wastewater. Furthermore, trickling filters normally do not slough at a uniform rate, adding additional day-to/day variability to unsettled effluent quality. All these factors are variables that need to be included in a rigorous conceptual approach to mathematically describing combined TF/AS performance. Although developing mathematical expressions to describe the performance of the suspended growth system is reasonably straightforward, the equations require a series of input parameters for which reliable data are normally unavailable. Suffice it to say that
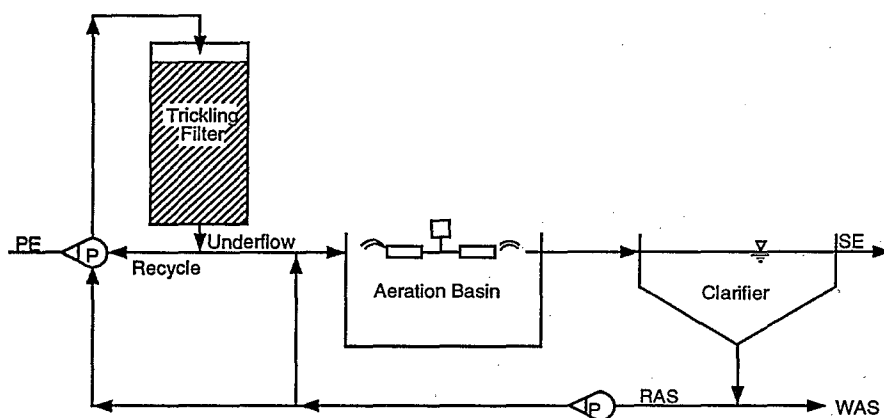

Figure 6-34. ABF/AS process schematic (adapted from Reference 148).

Table 6-26. Trickling Filter Removals at Livermore, California (Adapted from Reference 149)

	Filter Loading		
		lb BOD₅/	BOD₅
	kg BOD₅/m³/d	1,000 cu ft/d	Removal[a] (%)
Year			
1968–69[b]	1.92	120	32
1970	0.88	55	29
1971	0.83	52	28
1972	1.00	62	15
1973	0.82	51	18
1974	0.90	56	32

[a] Based on unsettled trickling filter effluent.
[b] During operation of one filter.

design of coupled TF/AS systems involves considerable empiricism.

A conservative design approach for a TF/AS system would be to ignore the BOD removal (influent minus unsettled effluent) occurring in the trickling filter, and base the activated sludge system design on the influent wastewater characteristics prior to the trickling filter. For roughing filter applications, where no nitrification occurs in the trickling filter, this approach has no impact on the minimum solids residence time, θ_c^m, required to sustain nitrification in the activated sludge system. It will result in overestimation of the net solids production in the AS reactor (for any θ_c) due to some active biomass decay in the TF, and an overestimate of oxygen requirements due to the combined impact of organic carbon oxidation and active biomass decay in the TF. One method of estimating these impacts has been presented (149). An alternative is to reduce biomass production and oxygen demand in the activated sludge unit by the biomass decay and estimated removals (based on unsettled effluent), respectively, which will occur in the trickling filter.

6.6.2.2 Nitrifying Trickling Filter/Activated Sludge and Nitrifying Trickling Filter/Solids Contact

When the trickling filter is designed to provide the level of nitrification required (Section 6.5.2.1), design of a coupled trickling filter/solids contact system is reasonably straightforward. Hydraulic retention times in solids contact vary from 3 to 60 minutes; a typical range of solids residence times is 0.2 to 1.0 day (152). Others indicate a general approach is to maintain θ_c between 1 and 2 days (153). Total organic loadings in TF/SC studies at Chino Basin and Garland were 0.2 to 1.1 kg BOD/m³/d (12 to 69 lb BOD₅/1,000 cu ft/d); for both studies, effluent TSS decreased with increasing θ_c up to between 0.5 and 1.0 days and were essentially constant at θ_c equal to 1.0 day (150). When short solids contact times are used, a sludge

reaeration tank is almost always needed to aerate solids adequately for proper bioflocculation. Where soluble BOD levels are low, as they will be for a nitrifying trickling filter, solids reaeration, either in a separate tank or by using step feed, is usually the preferred approach. Studies at Tolleson and Medford indicate that most flocculation occurs in less than 12 minutes (154). Secondary clarifiers are typically flocculator clarifiers. The combination of flocculator clarifiers in deep secondary tanks is reported to allow a monthly average effluent SS level of approximately of 10 mg/L at average overflow rates of 48 m³/m²/d (1,200 gpd/sq ft) (152).

It is also possible to have partial nitrification in the trickling filter and design for an additional increment of nitrification in the solids contact or activated sludge reactor. One conceptual approach is to design the SC or AS system with a θ_c^d that is $\leq \theta_c^m$ (or to use a process design factor near 1.0) and rely on the continual sloughing of nitrifying organisms from the trickling filter to sustain nitrification in the suspended growth reactor. For example, the theoretical relationship shown on Figure 6-35 illustrates the additional nitrification that can be achieved even when the design solids residence time, θ_c^d, is less than θ_c^m (the safety factor is less than 1.0) (155). This relationship assumes that nitrifiers continuously slough from the trickling filter at a uniform rate, dictated by the level of nitrifi-
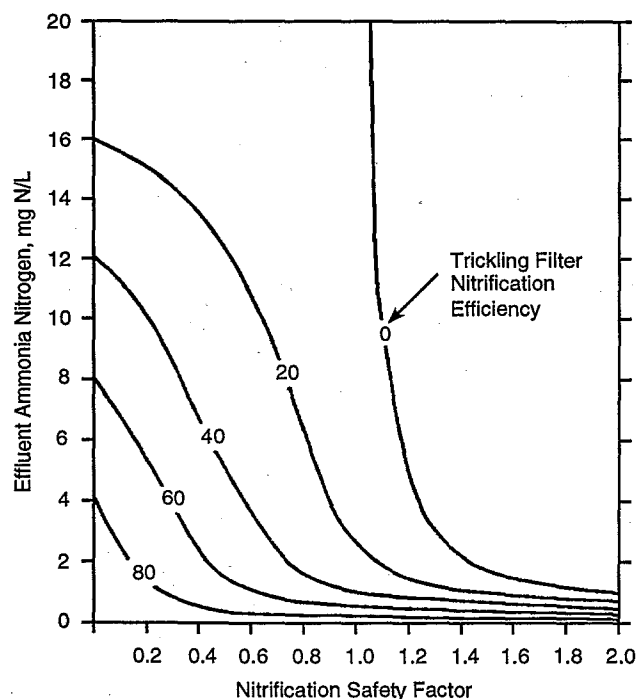
Figure 6-35. Effect of upstream nitrification on the effluent NH₄-N from downstream suspended growth unit. Influent NH₄-N is 20 mg/L and temperature is 20°C (adapted from Reference 155).

199

cation efficiency in the trickling filter. If the discharge of nitrifying organisms from sloughing fluctuates widely, the relationship in Figure 6-35 will not hold. Nitrification in the contact tank at Garland reportedly was inconsistent on a month-to-month basis indicating erratic seeding from the nitrifying trickling filter (156).

When the modeling approach used to develop Figure 6-35 was applied to data from the Duck Creek TF/SC plant, the results shown in Figure 6-36 were obtained. The Duck Creek plant consists of both solids reaeration, which is one-third of the total aerated volume, and solids contact; actual contact is 24 to 44 minutes. Even where the total solids residence time in the aeration basins was less than θ_c^m, additional nitrification occurred in the solids contact tank, indicating a reasonably constant degree of nitrifier sloughing from the TF. From 1988 to 1990, the TF loading averaged a moderate 0.4 kg/m^3/d (25 lb BOD$_5$/1,000 cu ft/d).

6.6.2.3 Aerated Biological Filter/Activated Sludge

The concept of returning activated sludge to an ABF was developed in 1966 by a wastewater treatment plant operator (157). In 1973, the original concept was modified by combining an aeration basin with the aerated biofilter (ABF/AS). General design and performance characteristics of the ABF/AS process have been reported (158). The process is currently marketed by Infilco Degremont.

As shown in Figure 6-34, recycle solids are mixed with the primary effluent and pumped through the tower. This approach provides a high initial F/M to the recycled solids which should contribute to improved sludge settleability.

With this arrangement, it is not possible to take any direct measurements of BOD$_5$ or COD reduction across the tower. Consequently, the contribution of the biotower to organic removals can be determined only indirectly. One approach to determining biotower contribution to satisfying the total oxygen demand is to measure the oxygen uptake rate in the aeration basin and compare this rate to the total COD reduction (influent COD minus effluent and waste COD) across the combined system (including additions for nitrogenous demand). Using a similar approach, it was estimated that the biotower might satisfy from 17 to 33 percent of the total oxygen demand (159).

Neptune Microfloc, which formerly offered the ABF/AS technology, summarized data from three pilot plant studies where nitrification was occurring (160). These data are presented in Table 6-27.

6.6.3 Other Inert Support Media

A variety of systems include addition of an inert support media to an activated sludge aeration basin (161). These media include freely moving porous pads (e.g., Captor and Linpor), trickling filter media (e.g., FAST and Bio-2-Sludge), racks of open weave media fixed in place (e.g., Ringlace), and modular media systems (Monitor). Undoubtedly, new approaches will continue to appear and some existing systems may not survive the competitive pressures of the marketplace. None of these systems is in widespread use now. The systems have certain process advantages but the cost benefits associated with their implementation are still being defined. The intent of this section is to provide a brief overview of some available systems.
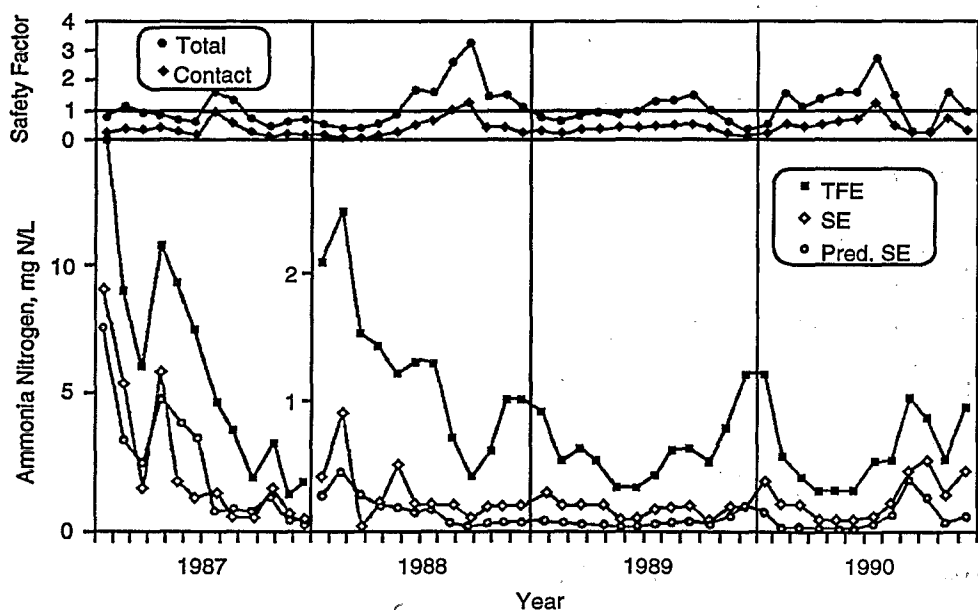

Figure 6-36. Comparison of predicted and actual monthly average effluent NH$_4$-N for Duck Creek (adapted from Reference 155).

200

Table 6-27. Nitrification Performance from ABS/AS Pilot Studies (Adapted from Reference 160)

Location	Influent (mg/L)			Effluent (mg/L)			Bio Cell	Aeration Basin		
	BOD	TSS	NH_4^+-N	BOD	TSS	NH_4^+-N	kg BOD_5/ m^3/d	MLVSS (mg/L)	Temp. (°C)	HRT (hr)
Corvallis,	56	64	15.0	20	29	2.2	0.69	1,690	11	3.8
Oregon	92	122	16.3	8	29	1.5	1.64	2,870	17	2.6
	113	153	16.0	6	23	0.9	1.35	2,350	20	4.1
	140	296	36.0	13	37	1.8	1.70	2,735	22	3.9
	198	296	46.0	15	51	3.4	5.62	2,820	23	3.4
Rochester,	286	181	25.0	12	27	1.1	3.43	3,210	17	8.3
Minnesota	221	174	24.8	8	18	0.9	2.66	3,580	17	7.1
	180	122	21.1	13	16	2.2	2.76	3,070	19	4.4
	167	101	12.7	14	5	0.7	2.65	3,625	23	3.7
Bend, Oregon	167	122	13.3	8	16	9.5	2.49	3,010	10	8.7
	186	198	17.4	11	30	5.1	4.33	2,810	13	5.4
	148	173	13.7	6	19	0.5	2.97	2,680	13	6.2
	141	141	11.2	5	18	0.1	3.53	2,420	15	5.1
	162	204	11.9	11	25	1.5	4.17	2,780	15	3.4
	184	207	12.3	12	27	0.5	3.85	2,870	18	4.1

6.6.3.1 Captor and Linpor

Both Captor (Ashbrook-Simon-Hartley) and Linpor (Linde AG) systems use porous pads freely suspended in the aeration basin. Linpor sponges are approximately cubical with sides of about 10 to 12 mm (0.4 to 0.5 in) and Captor sponges are about 12 by 25 by 25 mm (0.5 by 1 by 1 in). While not required, either system may include recycle of settled solids from the final settler back through the pads. The recycle option leads to a combined attached/suspended growth system. The pads are retained in the aeration basin via fine screens. The Captor system may include an airlift pump and pad conveyor system for squeezing excess solids from the porous pads. At one time, it was believed that this cleaning procedure would control solids loss from the pads in secondary treatment applications, such that existing aeration basins could be upgraded to nitrification by confining the pads to the upstream reactor segment where the bulk of the BOD and SS removal would occur. This upgrade would allow for a conventional activated sludge nitrification system in the downstream reactor segment. However, the ability to control solids loss never met expectations (162). The Linpor approach for combined carbonaceous removal and nitrification recycles settled secondary solids back through the pads. When treating secondary effluent in a separate stage nitrification application, there is normally no solids recycle.

Captor has been installed in Moundsville, West Virginia. Primary effluent is fed to a Captor zone, with 40 to 70

minutes hydraulic residence time which is directly followed by a conventional activated sludge reactor. Partial nitrification is achieved in the Captor zone. The only other full-scale Captor installation in the United States is in Opelousos, Louisiana, where Captor nitrifies secondary effluent, and is not followed by a final clarifier. One interesting application of Captor was a pilot plant investigation of lagoon effluent at Vesper, Wisconsin (163). The lagoon effluent was fed to a Captor reactor with no final clarifier. The ability to nitrify 0.1 to 0.16 kg NH_4^+-N/m3/d (6 to 10 lb/1,000 cu ft/d) at wastewater temperatures near 0°C (32°F) was demonstrated.

For combined carbonaceous removal and nitrification, the volume of pads in a Linpor system equals roughly 10 to 30 percent of the aeration tank volume. The fixed film biomass typically has an equivalent MLSS concentration of 1,200 to 3,800 mg/L and accounts for 25 to 60 percent of the total biomass in a typical aeration system (161).

The Freising, Germany, plant was converted to a Linpor system in 1984 by the addition of a pad volume equal to 20 percent of the aeration tank volume. The primary reasons for the conversion were frequently occurring poor sludge settleability, limited space, and cost effectiveness. Before conversion, the plant could only maintain an MLSS concentration of about 2,600 mg/L. Following conversion, a much higher MLSS concentration could be maintained and an average of 65-percent nitrification was achieved over a wastewater temperature range of 10°C (50°F) to 17°C (63°F), even though the DO concentration averaged

only 1.7 mg/L. Like Captor, the Linpor system has been employed as a fixed film system for nitrification of secondary effluent, where no final clarification or solids recycle occurred (164).

6.6.3.2 FAST and Bio-2-Sludge

Submerged trickling filter media systems include the FAST system (Smith and Loveless), Bio-2-Sludge (Weber Engineering), as well as other systems (165,166).

In a FAST system, typically 75 percent of the aeration tank volume is occupied by media. The media is 100 percent submerged and diffused aeration forces wastewater flow up through the media. A combined fixed film, suspended growth system was installed at Stow, Massachusetts, as illustrated in Figure 6-37. This plant was upgraded from a complete mix activated sludge system when new discharge standards limited total effluent nitrogen to 10 mg/L. The existing aeration tank was converted to a FAST system followed by an anoxic reactor equipped for methanol addition. The process design allows for either raw wastewater or methanol to be used as the carbon source. Methanol was used during startup; process performance during the first seven months is summarized in Table 6-28.

In contrast to FAST systems, the Bio-2-Sludge approach includes trickling filter media in only about 25 percent of the aeration tankage, but these systems always combine both attached and freely suspended (i.e., from sludge recycle) biomass.

Bio-2-Sludge systems use just about any synthetic media, as long as the openings are at least 2 by 2 cm (0.8 by 0.8 in). Typical media surface area ranges from 90 to 120 m^2/m^3 (27.5 to 36.6 sq ft/cu ft). Racks to hold the media are constructed to provide approximately 0.5 m (1.64 ft) of clear space between the top of the racks and the liquid surface. This arrangement allows for normal side roll mixing in the tank.

A combined carbonaceous oxidation-nitrification system using Bio-2-Sludge was installed in Schoemberg, Germany. Prior to plant conversion, the poor settling characteristics of the MLSS would not permit maintenance of more than 1,000 to 1,500 mg/L MLSS in the aeration tanks. After conversion and plant expansion, the MLSS settleability improved greatly (mean SVI of 82 mL/g) and secondary effluent quality averaged 0.4 mg/L NH_4^+-N (161).

Some investigators contend that in submerged biological filter systems with sludge recycle, the filter media is populated mainly by protozoa (166). The media then has only an indirect influence by making it easier to maintain a lower F/M ratio, due to the resulting improved SVIs.

6.6.3.3 Ringlace

In a Ringlace system (Ringlace Systems, Inc.), polyvinyl chloridene (PVCE) string on racks are installed in the
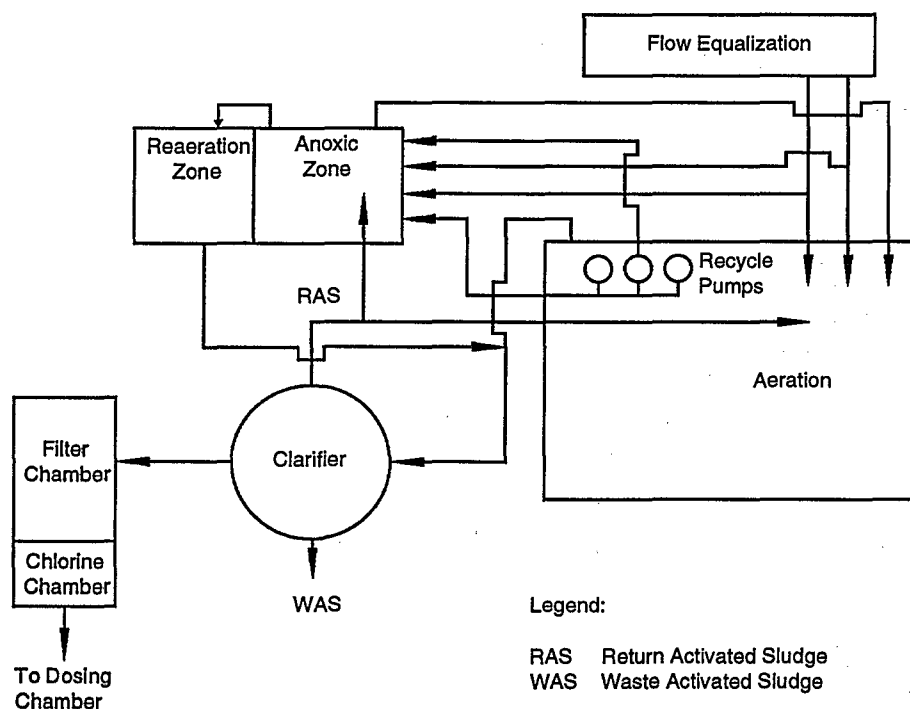

Figure 6-37. Nitrogen control schematic for Stow, Massachusetts (adapted from Reference 166).

Table 6-28. Average Influent and Effluent Data for Stow, Massachusetts (Adapted from Reference 167)

Month	BOD$_5$, mg/L		TSS, mg/L		NH$_4^+$-N, mg/L	
	Influent	Effluent	Influent	Effluent	Influent	Effluent
Mar.*	650	13	424	18	56	4
Apr.*	490	65	309	6	35	8.1
May	300	1	280	9	41	0.4
June	620	21	310	12	22	0.1
July	445	11	456	2	32	0.3
Aug.	230	3	311	1	38	0.1
Sept.	240	24	211	1	39	0.2

* Startup period.

aeration tank. Each string has numerous loops of the same material, thus greatly increasing the surface area available for fixed film growth. According to a European distributor, typically 25 to 50 percent of an aeration tank would contain Ringlace at a density of 120 to 300 lineal m/m^3 (11 to 28 ft/cu ft) of aeration tank (161). One U.S. distributor indicated the racks typically occupy about 50 to 80 percent of the basin volume with a mean density based on the entire basin volume of about 120 lineal m/m^3 (11 ft/cu ft).

These systems have been installed in Japan, with and without sludge recycle. The systems typically have been at small scale and lightly loaded. The Olching, Germany, plant added Ringlace in 1987. Before conversion, the maximum operational MLSS concentration achievable was about 1,500 mg/L, which resulted in a process F/M ratio of 0.6 to 0.7 kg BOD$_5$/kg MLSS/d. After conversion, the freely suspended biomass varied from 3,500 to 4,500 mg/L. Fixed biomass on the Ringlace material was estimated at 6.5 g/m which is equivalent to an additional 790 mg/L of MLSS. The overall operating F/M of the system decreased to approximately 0.2 and the desired levels of nitrification could be achieved (161). Ringlace Systems Inc. is scheduled to supply the media for a technology evaluation in one of the aeration basins at the 440 L/s (10 mgd) Annapolis Wastewater Reclamation Facility.

6.6.3.4 Monitor

The final system to be mentioned is Monitor (KLV Technologies)—a new system that consists of a media-filled biochamber constructed on pontoons and floated into position. This system is suitable for lagoon installations that require upgrading for nitrification. Wastewater is pumped into and through the aerated media. A Monitor system in Ontario County, New York, was reported by the manufacturer to maintain ammonia-nitrogen concentrations below 0.5 mg/L in cold wastewater temperatures (168).

6.6.3.5 Design Considerations

When fixed film and freely suspended biomass systems are combined, the overall contribution from each system component typically is uncertain. Because of mass transfer limitations for both substrate and DO, a unit of attached biomass usually is not equivalent to a unit of freely suspended biomass in terms of ability to oxidize carbon or ammonia. Since some media systems are more porous or open than others, the benefits of alternative media systems should not be compared solely on the mass of biomass attached. Furthermore, the advantages frequently associated with such systems (e.g., improved settleability) cannot be related to a specific design parameter, such as specific surface area available per unit of loading. Given current levels of understanding, pilot plant evaluations remain the best design approach whenever these systems are under consideration.

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National Technical Information Service
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Springfield, VA 22161
703-487-4650

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Chapter 7
Design Considerations for Denitrification Processes with Supplemental Substrate Addition

7.1 Introduction

As discussed in Chapter 2, the process of biological denitrification can be used to remove nitrogen from wastewater when the nitrogen is predominantly in the form of nitrate. In municipal applications, nitrogen is present in raw wastewater primarily in organic and ammonium-nitrogen form and must be converted (nitrified) to an oxidized form (nitrite or nitrate) before biological denitrification can take place. Nitrification and denitrification can occur separately in distinct systems (Chapter 6) or in one system where the process would be termed single-sludge (Chapter 8). Basic chemistry, microbiology, and kinetics are all relevant to nitrification and denitrification (Chapters 3 and 4, respectively).

This chapter presents design criteria for alternative denitrification systems, including suspended growth and attached growth processes, that use a supplemental carbon source. The attached growth processes that are discussed include the downflow packed-bed and upflow fluidized-bed systems. Other systems have been investigated but are not currently in full-scale use (1).

7.1.1 Classification of the Supplemental Substrate Addition Denitrification Process

The supplemental substrate addition process generally involves a separate-stage denitrification process, using a distinct biological process to remove nitrate. Nitrification is accomplished prior to the denitrification step in either a combined carbon-oxidation and nitrification process or in a separate system.

A combined carbon-oxidation and nitrification system with a separate denitrification system is known as a two-sludge system. When oxidation, nitrification, and denitrification occur separately, the process is called a three-sludge system. Regardless of whether oxidation and nitrification are combined, a separate denitrification process is used for both the two-sludge and three-sludge processes. The addition of supplemental substrates to single-sludge systems is addressed in Chapter 8.

7.1.2 Substrate Selection

A supplemental carbon source is needed in separate-stage denitrification systems because the preceding carbon-oxidation and nitrification step has removed nearly all of the degradable carbonaceous material from the wastewater. Several alternative sources of carbon have been considered at various plants; however, only a few of these, such as methanol, acetate, and raw sewage, have actually been tested. Alternative sources include (2,3,4):

- Methanol

- Raw sewage or primary effluent

- Ethanol

- Methane

- Acetate

- Ketones

- Molasses and sugars

- Brewing and distilling wastes

The ideal supplemental substrate should be inexpensive, readily available all year, essentially free of nitrogen (as well as phosphorus, in many cases), and should have suitable kinetics. (Substrate selection is discussed in Chapter 4.) Neither raw sewage nor primary effluent is generally suitable as a carbon source because of high ammonia, organic nitrogen, and SS levels; however, if the introduced nitrogen levels can be tolerated, these substrates may be acceptable. In most cases, methanol is the most appropriate choice and is being used in the vast majority of installations because of its availability, low cost, favorable sludge production, low volatile organic compound (VOC) emissions potential, and lack of nitrogen and phosphorus. Because of its near universal use, methanol is the only supplemental substrate considered in this chapter.

The methanol dose should be monitored and controlled carefully to prevent overdosing, which can adversely affect the effluent BOD; however, some systems are reported to work well with little attention to monitoring the

methanol dose. The methanol dose is based on the nitrogen removal required; the methanol to nitrogen ratio (M:N) is typically 2.5-3.0. (See Chapter 4 for discussion of the stoichiometry.) Details regarding methanol properties and handling are presented at the end of this chapter.

7.2 Suspended Growth Systems

7.2.1 Description

Suspended growth denitrification is an activated sludge process. The biomass is kept in suspension in a reactor by mixing before being allowed to settle out in a clarifier, and the majority of the settled biomass is recycled to the reactor. In the standard activated sludge process, oxygen is supplied to act as an electron acceptor for the oxidation of aerobic carbonaceous matter in wastewater. In the denitrification process, nitrate acts as an electron acceptor and methanol is supplied to the reactor to serve as the carbonaceous matter. The continuous flow-through reactors are typically operated with an average detention time of 2-3 hr (5). An aeration basin or zone must be provided after the denitrification reactor to strip the nitrogen gas bubbles produced during the denitrification process and oxidize any methanol that remains. Nitrogen gas bubbles must be removed prior to the final clarifier to prevent problems with floating sludge caused by adherence of the nitrogen bubbles to the sludge. A portion of the settled sludge is returned to the basin and the remainder is wasted to maintain the desired solids residence time (θ_c). A typical suspended growth denitrification system with supplemental substrate addition is shown schematically in Figure 7-1.

A suspended growth system with methanol addition has the following reported or potential advantages:

- Because the rate of denitrification should be greater than if wastewater were used as the carbon source, reactors can be smaller.

- Excess methanol is oxidized more easily than in attached growth systems.

- Because nitrification and denitrification occur in different stages, each process can be optimized separately.

- A high degree of nitrogen removal is possible.

- A plant retrofit can likely be accomodated.

- Suspended growth processes are well-understood by most operators although experience with separate-stage suspended growth systems for denitirification is limited.

A separate-stage suspended growth system with methanol addition has the following disadvantages:

- Methanol is an added expense.

- A separate clarification step is required.

- A greater number of unit processes is required for total nitrogen removal when compared to a single-sludge system.

7.2.2 Design Considerations

The design considerations for suspended growth denitrification systems are similar to the requirements for standard activated sludge plants. Key factors are hydraulic residence time (HRT), solids residence time (θ_c), and methanol requirements. Systems are designed on the basis of reaction kinetics and full- and pilot-scale experience. Chapter 4 discusses the basic kinetics of suspended growth denitrification. Table 4-1 summarizes some of the kinetic coefficients that have been determined for suspended growth systems where methanol is used as the carbon source. There is significant variation in the data presented in this table, due to differences in test conditions and procedures. Care must be taken if these data are used without consulting the specific reference.

The aerobic reactor that follows the anoxic reactor (to strip out nitrogen gas bubbles and oxidize excess methanol) should be mildly aerated so the floc does not break up and sedimentation is not hindered (6). The detention time of the aerobic reactor is typically 20–60 min for ni-
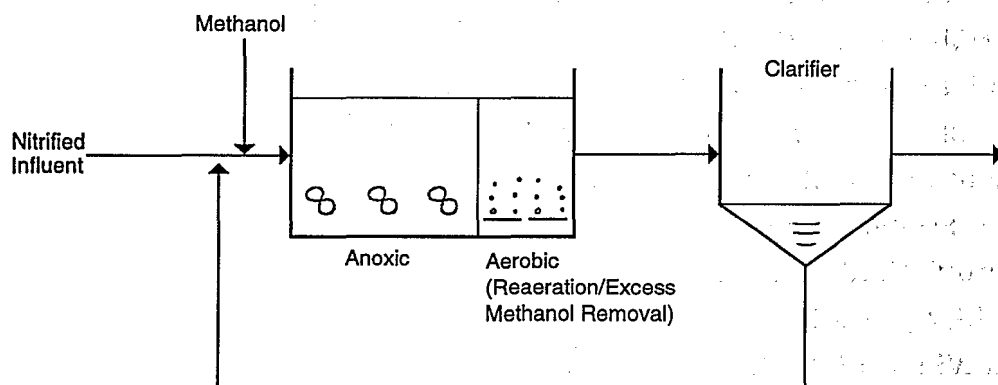

Figure 7-1. Schematic of suspended growth system.

trogen gas stripping and excess methanol oxidation. Requirements for oxidizing excess methanol are further discussed in Section 7.5.7; kinetic equations for methanol oxidation are generally used to estimate the required reactor residence time.

The design considerations for solids-liquid separation (clarification) for suspended growth denitrification systems are the same as those for nitrification and standard activated sludge systems (see Chapter 2). This section will not attempt to discuss aspects of clarifier design. The reader is directed to WPCF FD-8 (7) and WEF MOP 8 (8) for a more detailed discussion.

7.2.3 Design Example

The design of suspended growth denitrifying reactors with methanol addition can be based on the kinetic equations presented in Chapter 4. The design procedure presented here is based on those equations, although other protocols can be followed. The equations and the kinetic coefficients that are used are intended for illustration purposes only.

A design example for a suspended growth activated sludge system receiving a nitrified effluent follows. In an actual design situation, it is necessary to develop a flow and materials balance such as generated for the more complex Plant B shown in Figure 2-6 and illustrated in Table 2-16. Assume such an analysis yields the nitrified

effluent criteria summarized in Table 7-1. The effluent criteria are for year-round total nitrogen limits and the more stringent BOD and TSS limits shown in Table 2-10. Table 7-1 also lists the effluent limits and the effluent values used for this design example.

Table 7-1. Design Example: More Complex Plant B with Year-Round Effluent Limits—Suspended Growth System

Characteristic	Nitrified Effluent	Effluent Limits	Design Final Effluent
Minimum Monthly Temp.	—	—	15°C
Average Flow, m³/d	18,930	—	—
Average Flow, mgd	5	—	—
TSS, mg/L	15	10	8
CBOD$_5$, mg/L	3	10	8
TKN, mg/L	0.7	—	—
$(NO_3^- + NO_2^-)$-N, mg/L	24.5	—	—
TN, mg/L	25.2	5	3

Note: Design final effluent values are lower than the effluent limits for reliability.

7.2.3.1 Design Example: Suspended Growth Denitrification System

1. Select the kinetic coefficients and other design constants to be used. The selection of the kinetic coefficients can be critical to the design. The values in Table 4-1 may be used as a starting point. Since there is significant variation in the values reported in the literature, care must be taken when selecting values, and consideration should be given to conducting pilot studies. When considering the need to conduct a pilot study, one must weigh its cost against the potential savings that may result from the use of less conservative kinetic-coefficients that might be developed from the pilot evaluation. The studies would also provide site-specific information that might allow the application of lower design (or safety) factors.

The following design constants have been selected for purposes of illustration:

- True (gross) yield: Y_s (g VSS/g COD removed) = 0.18
- Decay coefficient: b_d (g VSS/g MLVSS/d) = 0.04
- Maximum substrate removal rate: $\hat{q}_s$ (g COD/g VSS/d) = 10.3
- Half-saturation constant: K_s (mg/L COD) = 9.1
- Temperature correction coefficient: θ = 1.08
- Maximum month MLSS, mg/L = 3,000
- Design MLSS, mg/L = 2,000
- Assumed MLVSS/MLSS = 0.75

7.2.3.1 Design Example (continued)

2. Estimate the amount of methanol allowed in the effluent. Start by estimating the effluent BOD_5 that will come from the solids in the effluent.

- Assume BOD_5 of the effluent VSS = 0.5 mg BOD_5/mg VSS (see Chapter 2)
- Assume VSS of the effluent TSS = 0.75 mg VSS/mg TSS (see Chapter 2)
- Thus, effluent BOD_5 from TSS = 8 x 0.75 x 0.5 = 3 mg/L

Estimate the amount of BOD_5 allowed in the effluent that may be from excess methanol. Assume that the BOD_5 of the secondary effluent will pass through the denitrification system unchanged. Thus, the maximum amount of BOD_5 in the effluent because of excess methanol may be estimated as:

- Allowable BOD_5 from methanol = design effluent BOD_5 (Table 7-1) − secondary effluent BOD_5 (Table 7-1) − BOD_5 from TSS = 8 mg/L − 3 mg/L − 3 mg/L = 2 mg/L
- Convert this BOD_5 to a COD value based on a BOD_5:COD ratio of 0.57 for methanol (9)
- Allowable COD from methanol = 2 mg/L BOD_5 ÷ 0.57 = 3.5 mg/L COD

This is the maximum amount of COD allowed in the effluent because of excess methanol. The volume of the anoxic zone can be reduced by allowing the methanol concentration in the anoxic reactor to be greater than this amount, thus increasing the rate of denitrification. The excess methanol could be removed in the subsequent aerobic reactor. There will be a tradeoff between the costs for supplying this excess methanol and the capital costs of the larger anoxic reactor that would be required if the methanol concentration were to be kept lower. For this example, a methanol concentration of about half the saturation constant will be used, yielding a COD of 4.5 mg/L in the anoxic reactor. This is equivalent to 3 mg/L of methanol, based on the stoichiometric COD:methanol ratio of 1.5.

3. Estimate what the concentration of nitrate in the effluent must be to achieve the design effluent total nitrogen concentration. First, assume that the soluble, nonbiodegradable organic nitrogen in the effluent is 1 mg N/L. Second, assume that the TKN:VSS ratio of the solids in the effluent is 0.08 mg N/mg VSS.

- Effluent particulate N = effluent TSS x VSS:TSS x 0.08 = 8 x 0.75 x 0.08 = 0.48 mg N/L
- Thus, total organic N = 1 mg N/L + 0.48 mg N/L = 1.48 mg N/L
- Based on the calculations in Chapter 6, it can be assumed that the ammonia concentration in the nitrified effluent is 0.5 mg N/L
- Thus, the total effluent TKN (ammonium plus organic nitrogen) anticipated from the denitrification reactor = 0.5 + 1.48 = 1.98 mg N/L, say 2.0 mg N/L
- Thus, the allowable nitrate in the effluent = design effluent TN − TKN = 3 mg N/L − 2.0 mg N/L = 1.0 mg N/L
- Base the design on an effluent nitrate of 1.0 mg N/L
- Thus, the average nitrate removal = nitrified effluent nitrate − design effluent nitrate = 24.5 mg N/L − 1.0 mg N/L = 23.5 mg N/L

4. Calculate substrate removal rate. Equations based on organic substrate removal (on a COD basis) are used for this example. Assuming that the methanol will be the only rate-limiting factor,

$$q_s = \hat{q}_s \left(\frac{S}{K_s + S} \right) = (10.3) \left(\frac{4.5}{9.1 + 4.5} \right) = 3.41 \text{ g COD/g VSS/d}$$

- This rate can be related to an equivalent nitrate removal rate:

$$q_D = (1 - 1.42 \, Y_s) \frac{q_s}{2.86} + \frac{1.42}{2.86} b_d$$

214

7.2.3.1 Design Example (continued)

$$q_D = (1 - 1.42(0.18)) \frac{3.41}{2.86} + \frac{1.42}{2.86} \, 0.04$$

$$= 0.91 \text{ g NO}_3^-\text{-N/g VSS/d}$$

Based on the nitrate removal rates presented in Table 4-1, this value appears reasonable.

- Convert to a rate at the minimum monthly temperature:

$$q_{s_{15}} = q_{s_{20}} \, \theta^{(15-20)} = 3.41 \, (1.08)^{-5}$$

$$= 2.32 \text{ g COD/g VSS/d}$$

5. Calculate minimum required solids residence time (θ_c^m) based on the above rate:

$$\frac{1}{\theta_c^m} = Y_s q_{s_{15}} - b_d$$

$$= (0.18)(2.32) - 0.04 = 0.38 \text{ d}^{-1}$$

$$\theta_c^m = 2.6 \text{ d}$$

6. Select the peaking factor for design. From Table 2-18, a total process peaking factor of 1.56 was selected.

7. Calculate the design SRT (θ_c^d):

- $\theta_c^d = \theta_c^m \times$ design peaking factor = 2.6 d x 1.56 = 4.1 d

8. Calculate methanol dose required for denitrification from Equation 4-10. Assume the nitrified effluent DO is 2 mg/L.

- Methanol dose = 2.47 x (nitrate removed) + 0.87 x (DO) + design methanol in anoxic reactor = 2.47 x 23.5 mg N/L + 0.87 x 2 mg/L + 3 mg/L = 63 mg methanol/L

- Convert to COD: 63 mg/L x 1.5 = 94.5 mg COD/L, or 1,790 kg (3,940 lb) COD/d

9. Calculate required anoxic HRT and reactor volume. For simplicity, assume that the incoming VSS in the nitrified effluent do not decay further in the denitrification reactor. Also the small residual effluent methanol COD concentration can be ignored for a slightly more conservative design.

$$\text{HRT} = \frac{Y_s \, (S_o - S) \, \theta_c^d}{(1 + b\theta_c^d) \, \text{MLVSS}} + \frac{(\text{VSS}) \, (\text{SRT})}{\text{MLVSS}}$$

$$= \frac{0.18 \, (94.5 - 0) \, (4.1)}{(1 + (0.04) \, (4.1) \,) \, (2,000) \, (0.75)} + \frac{(15) \, (4.1) \, (0.75)}{2,000 \, (0.75)}$$

$$= 0.04 + 0.031 = 0.071 \text{ d}$$

$$= 1.704 \text{ hr}$$

Use a design HRT of 1.75 hr.

- Calculate the volume of the reaction/postaeration tank, assuming a 40-min HRT to provide adequate stripping of nitrogen gas and removal of the excess methanol

- The total reactor HRT is 2.4 hr, 73% anoxic and 27% aerobic:

    Volume = 18,930 m³/d x 2.4 hr x (1 d/24 hr) = 1,893 m³ (500,000 gal)

7.2.3.1 Design Example (continued)

- Use three reactors, 4.6 m (15 ft) deep with a length:width ratio of 3:1:

    Reactor Volume = 1,893 m³/(3 reactors)

    = 631 m³ (22,280 cu ft) each

    Reactor Volume = (3W) (W) (Depth)

    631 m³ = (3W²) (4.6)

    Width = 6.8 m (22.3 ft)

    Length = 3 x 6.8 m = 20.4 m (67 ft)

Thus, each of 3 reactors is 20.4 m x 6.8 m x 4.6 m (67 ft x 22.3 ft x 15 ft) deep. Allow approximately 0.6 m (2 ft) freeboard in final tank sizing (total depth of 5.2 m [17 ft]).

Compartmentalizing the reactor to promote plug flow characteristics will ensure the best performance.

10. Calculate the size of the denitrification clarifiers. Check sizing based on a surface overflow rate (SOR) and a solids loading rate (SLR) and select the larger of the two. Use maximum MLSS of 3,000 mg/L.

- Assume a design SOR of 20.4 m³/m²/d (500 gpd/sq ft) at average conditions

- Area required for SOR = 18,930 m³/d ÷ 20.4 m³/m² = 928 m² (10,000 sq ft)

- Assume a design SLR of 122.3 kg/m²/d (25 lb/d/sq ft) at the maximum month MLSS concentration of 3,000 mg/L (3 kg/m³). Assume a return sludge rate of 0.6 the incoming flow

- Size required = (1 + 0.6) x 18,930 m³/d x 3 kg/m³ ÷ 122.3 kg/d/m² = 743 m² (8,000 sq ft)

- Thus, SOR controls the design sizing. Use three clarifiers to provide adequate surface area:

    (928 m²)/3 clarifiers = 309 m²/clarifier

    $\pi D^2/4$ = 309 m²

    D = 19.9 m (65 ft)

Use three 19.9-m (65-ft) diameter clarifiers.

11. Calculate the horsepower required for mixing the denitrification reactor:

- Assume 13.2 hp/1,000 m³ (50 hp/Mgal) (see Chapter 2)

    Anoxic Basin Volume = 1,893 m³ x 73% anoxic = 1,382 m³ (370,000 gal)

    hp = (1,382) (13.2/1,000) = 18.5 hp/3 basins, or 6.2 hp/basin

12. Calculate rate of waste sludge production.

    Net Yield = $Y_{S_{NET}}$ = Y_s / (1 + $b_d \theta_c^d$)

    $Y_{S_{NET}}$ = (0.18) / [(1 + 0.04 (4.1)]

    = 0.155 kg VSS/kg COD removed

    = (0.155 kg VSS/kg COD) / (0.75 kg VSS/kg TSS)

    = 0.21 kg TSS/kg COD removed

    Waste Sludge = Q (S_o – S) ($Y_{S_{NET}}$) + Q (Influent TSS removed)

    = [(18,930) (94.5 – 0) (0.21) + (18,930) (15 – 8)]/(1,000)

    = 376 + 133 = 509 kg (1,120 lb)/d

A process flow diagram is shown in Figure 7-2 for the separate-stage suspended growth denitrification system.
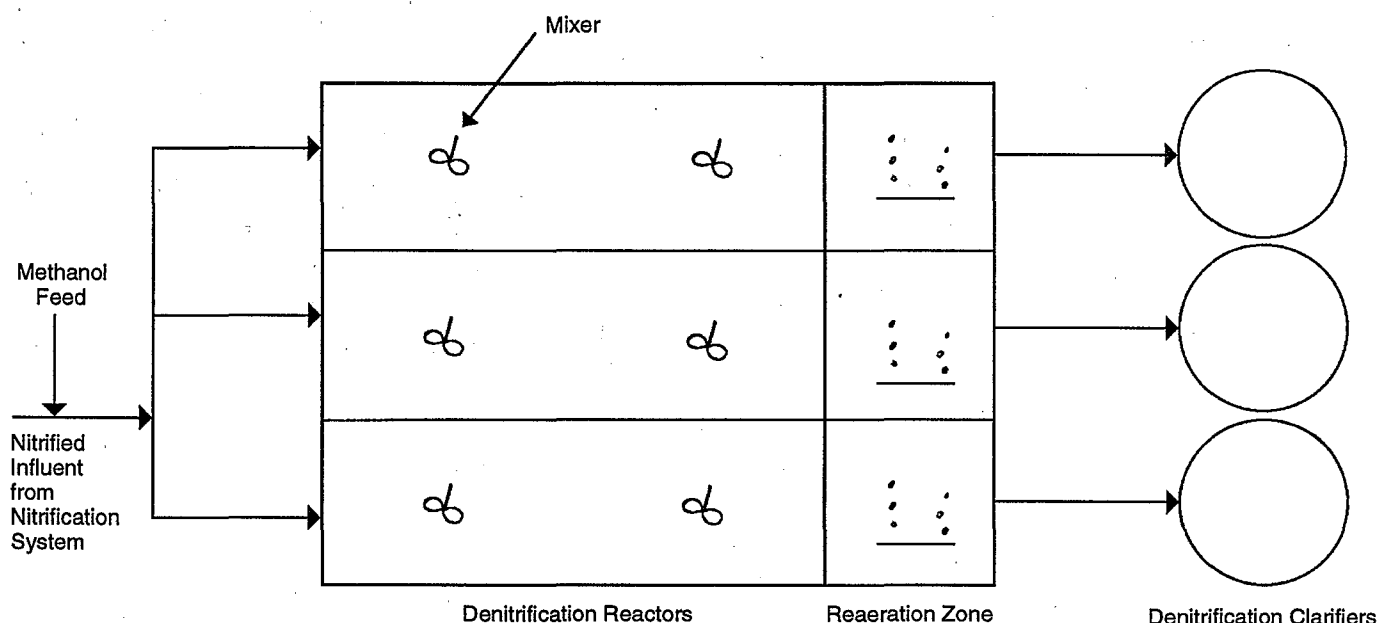
Figure 7-2. Design example schematic of suspended growth system.

7.2.3.2 Alternate Design Approach

As previously indicated, there are alternate design approaches to the one illustrated in Section 7.2.3.1. An alternative procedure would be to rely on rate data such as presented in Figure 7-3. At 15°C (59°F), denitrification rates of 0.05–0.18 kg NO_3^--N/kg MLVSS/d are shown. This can be compared to the rate for the design in Section 7.2.3.1, which equates to a removal rate of 0.21 kg NO_3^--N/kg MLVSS/d for the denitrification of 23.5 mg/L of NO_3^--N in the 1.75-hr detention time anoxic reactor, operated at an MLVSS of 1,500 mg/L. Using the data in Figure 7-3 as an alternate design basis would produce a larger anoxic reactor whose size would depend on the rate selected.

7.2.4 Case Study: River Oaks Advanced Wastewater Treatment Plant (Hillsborough County, Florida)

The River Oaks Advanced Wastewater Treatment Plant (AWTP) was upgraded in three phases beginning in 1986. The upgrades increased capacity from 131 L/s (3 mgd) to 438 L/s (10 mgd) and included the addition of suspended growth denitrification with methanol addition. The completed plant includes flow equalization, headworks, primary clarification, aeration, secondary clarification with final flocculation, suspended growth denitrification, and disinfection. A flow schematic is shown on Figure 7-4.

The denitrification system includes two 3,940-m³ (1.04-Mgal) tanks operated in parallel with 16 cells each. The cells are further divided into anoxic and aerobic zones. The first 10 cells operate in the anoxic mode; the next 4 can operate in either the anoxic or aerobic mode; and the last 2 cells operate only in the aerobic mode.

The plant discharges to Tampa Bay and was designed to meet the following stringent effluent limits: BOD_5 = 2 mg/L, TSS = 2 mg/L, TN = 1.2 mg/L, and TP = 0.4 mg/L, all based on a maximum monthly flow of 526 L/s (12 mgd). The limits have since been relaxed to 5, 5, 3 and 1 mg/L, for BOD_5, TSS, TN, and TP, respectively, on an annual average. Monthly plant operating data are shown in Table 7-2.

The River Oaks AWTP has demonstrated the ability to keep TN levels below 1.0 mg/L when operated at approximately 70 percent of design flow. The treatment processes employed have provided for flexible and reliable treatment and have demonstrated the ability to remove nitrogen to low levels. The methanol used for denitrification represents about 10 percent of the plant O&M costs.

7.3 Downflow Packed-Bed Systems

7.3.1 Description

The downflow packed-bed system is an attached growth process. Physically, it is identical to a deep-bed downflow sand filter. Denitrifying microorganisms attach to the filter media which provides the support system for their growth. Methanol is added upstream of the packed-bed filter and the nitrified effluent is filtered through the media. The packed-bed filter system is well suited for denitrification because it provides the necessary hydraulic detention time for the biological reaction to take place, and the open water surface in the filter is sufficiently limited to minimize transfer of oxygen by surface aeration. A schematic of a typical system is shown in Figure 7-5, and a cross-section schematic is shown in Figure 7-6. The packed-bed system is a proprietary process of TETRA Technologies,
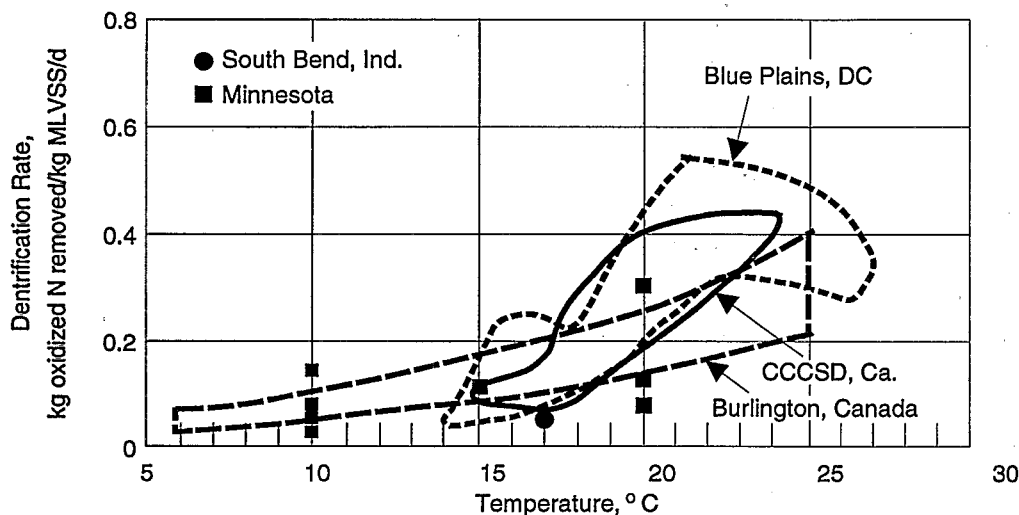
Figure 7-3. Observed denitrification rates for suspended growth systems using methanol (from Reference 10).
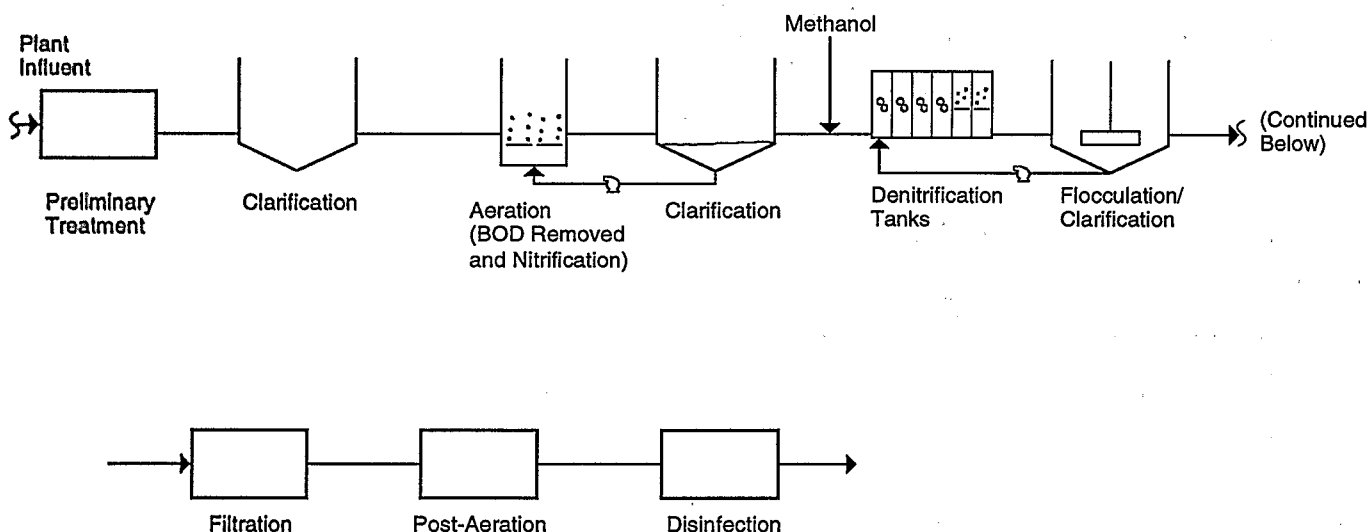

Figure 7-4. Schematic of River Oaks Advanced Wastewater Treatment Plant (from Reference 11).

which holds a patent based on a previous Dravo patent (12). A representative installation list is shown in Table 7-3.

The filter is composed of coarse, hard, round, high-density media. These media filter out solids and serve as a support system for the the denitrifying microorganisms. The packed-bed system also eliminates the requirement for downstream filtration or clarification.

7.3.2 Design Considerations

As denitrification occurs, nitrogen gas accumulates in the filter media, which increases the headloss. The nitrogen gas bubbles are periodically released from the media by taking the filter off line and applying backwash water for

a few seconds—a process commonly called the nitrogen release cycle (NRC) or filter bumping (13). The frequency of the NRC is a function of both nitrate removal and a minimum acceptable time between cycles, usually no less than 1 hr. Usually, a filter needs to be bumped once every 4–8 hr, depending on the nitrogen loading.

As with a conventional gravity filter, SS gradually accumulate in the filter, and part of the nitrogen removal will occur because nitrogen in the SS is removed. Physical-removal mechanisms may include straining, sedimentation, and particle interaction. Chemical-removal mechanisms may include adsorption, coagulation-flocculation, and biological activity. As with all filters, no one

218

Table 7-2. River Oaks Wastewater Treatment Plant: Monthly Performance Data

| Month	Plant Influent						Denitrification Process Influent					Denitrification Process Effluent					Final Plant Effluent					
	Flow		BOD$_5$	TSS	TKN	NH$_4^+$-N	BOD$_5$	TSS	TKN	NH$_4^+$-N	NO$_x$-N	BOD$_5$	TSS	TKN	NH$_4^+$-N	NO$_x$-N	BOD$_5$	TSS	TKN	NH$_4^+$-N	NO$_x$-N	TN
	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	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L
SEP 90	305	6.96	167	183	31.6	23.7	2.0	0.5	1.27	0.11	13.20	2.0	7.0	0.93	0.10	0.56	1.00	1.00	0.64	0.10	0.47	1.12
OCT90	316	7.21	174	156	34.0	23.4	2.0	0.7	1.59	0.12	11.37	2.0	6.0	0.42	0.10	0.07	1.00	0.00	0.73	0.10	0.10	0.80
NOV 90	292	6.66	212	173	34.9	24.5	3.0	0.8	1.47	0.13	16.30	3.0	9.0	1.04	0.10	0.33	1.00	1.00	0.71	0.10	0.30	1.01
DEC 90	308	7.03	237	200	36.5	28.0	3.0	0.8	1.56	0.14	15.90	4.0	11.0	2.40	0.10	0.56	1.00	1.00	0.74	0.10	0.58	1.33
JAN 91	329	7.54	215	185	33.4	27.3	3.0	0.1	6.00	2.39	13.32	2.0	6.0	4.62	2.35	0.71	1.00	1.00	3.08	2.28	0.72	3.80
FEB 91	334	7.62	194	133	35.6	19.6	5.6	10.7	2.65	0.55	13.36	4.2	8.0	1.23	0.15	0.76	1.00	0.60	0.88	0.17	0.83	1.71
MAR 91	317	7.23	213	167	38.7	26.2	3.6	7.4	2.09	0.26	14.79	4.3	8.1	1.36	0.11	0.48	0.83	0.55	0.72	0.10	0.55	1.21
APR 91	290	6.61	200	142	38.0	26.0	3.7	4.9	1.64	0.17	14.54	3.6	5.2	1.41	0.10	0.14	0.87	0.13	0.75	0.10	0.13	0.89
MAY 91	317	7.23	173	146	37.7	24.7	2.4	3.7	1.79	0.18	12.74	6.0	5.7	1.28	0.10	0.19	0.80	0.32	0.89	0.10	0.25	1.14
JUN 91	309	7.04	177	130	34.5	23.4	3.2	3.6	1.53	0.13	12.39	4.8	8.2	1.40	0.10	0.23	1.09	0.50	0.87	0.10	0.34	1.21
JUL 91	340	7.77	167	127	31.5	21.5	2.9	7.1	1.31	0.14	10.21	3.9	5.8	1.04	0.10	0.24	1.65	0.40	0.79	0.10	0.20	0.98
AVG.	314	7.17	194	158	35.1	24.4	3.1	3.7	2.08	0.39	13.47	3.6	7.3	1.56	0.31	0.39	1.02	0.59	0.98	0.30	0.41	1.38
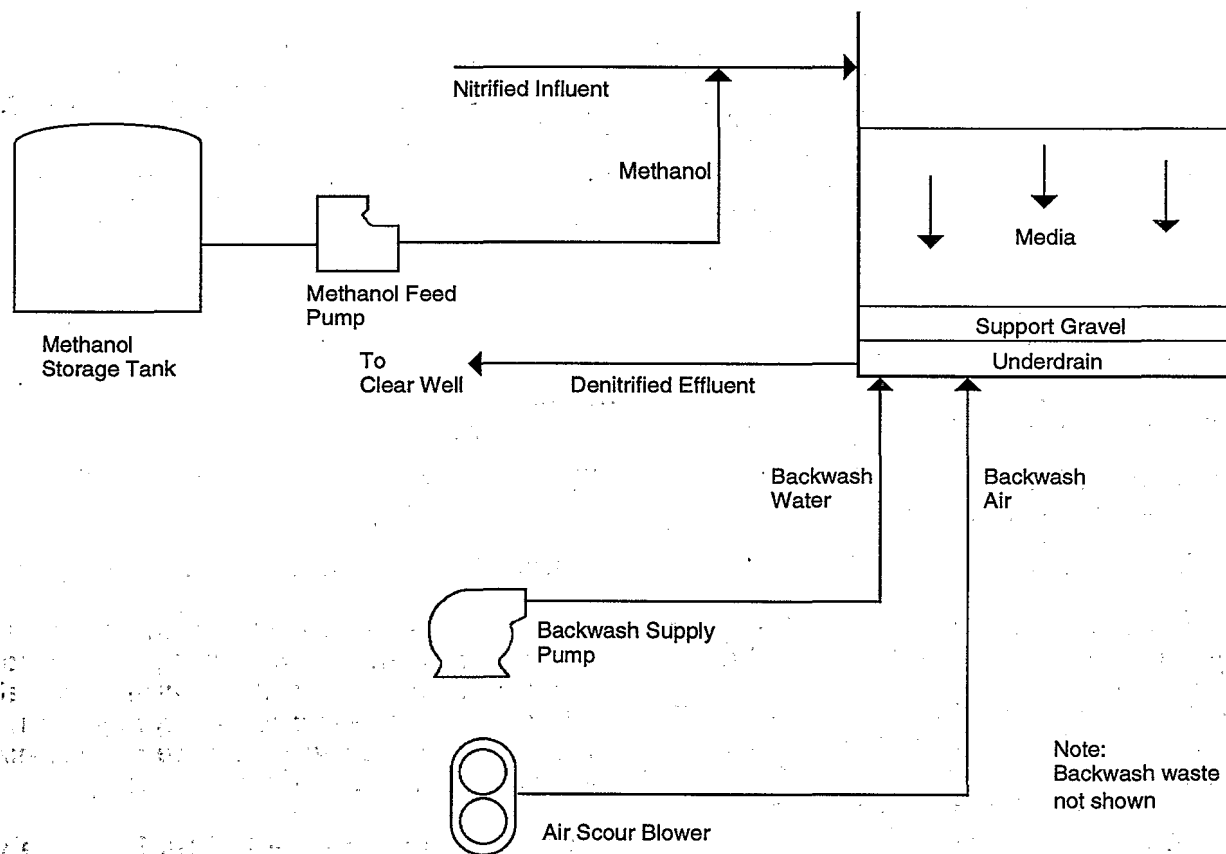
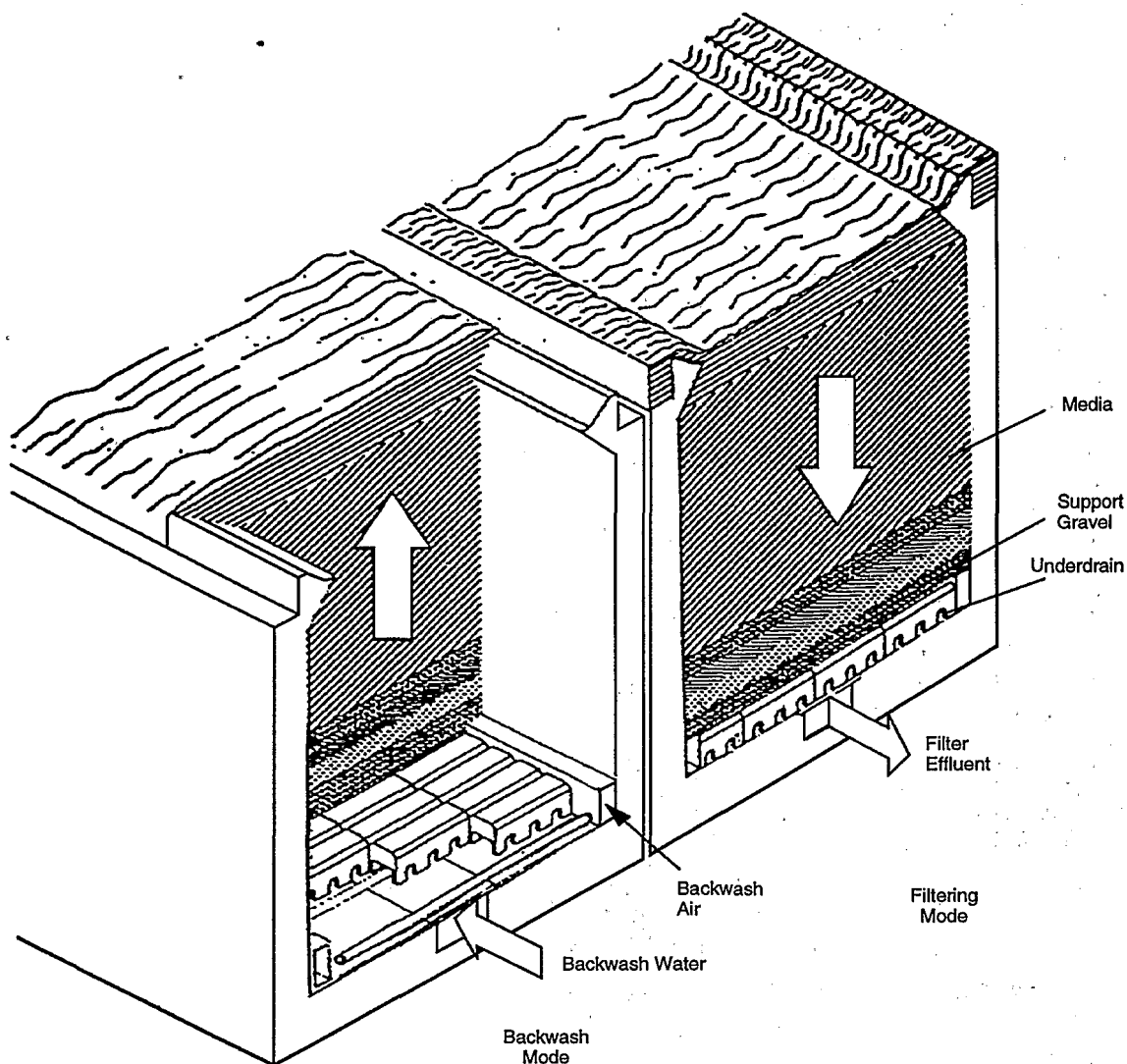Figure 7-5. Linear schematic of downflow packed-bed system.

Figure 7-6. Cross-section schematic of downflow packed-bed system.

mechanism can account for the total effect. The SS trapped in the filter are removed by backwashing, usually with an air/water wash in a step that also sloughs off excess biomass. Part of the backwash waste may be returned to the filter to reseed the filter with denitrififying microorganisms. The requirement for reseeding is dependent on the influent TSS, the efficiency of backwashing, and the presence of methanol-acclimated biomass. Therefore, it may not be desirable to completely clean the filter during backwash. Return flows equivalent to 880–1,170 m³/m²/d (15–20 gpm/sq ft) of filter for 1–2 min may be used if reseeding is needed. After backwashing, a loss of denitrification capability may be observed because of the loss of accumulated biomass. A cleaner filter placed in operation requires some time to reestablish the biomass and may experience some decreased performance when first placed back in service. Therefore, several small filters will produce a better blended effluent than a

few large filters. If the accumulation of SS were allowed to continue without backwashing, the filter would become clogged and the frequency of gas bumping would have to be increased. Under normal conditions, headlosses will require that the filter be automatically backwashed every 1–5 d, which is comparable to the backwashing frequency for a filter used only for SS removal.

The design of denitrification filters must consider filtration, denitrification kinetics, and the limitations on the frequency of the NRC. The process hydraulic loading rate is generally 58–117 m³/m²/d (1–2 gpm/sq ft) with an empty bed contact time of 30 min or greater. Figure 7-7 illustrates typical design curves based on empty bed contact time.

Knowledge of reaction rates is necessary to size an attached-growth denitrification reactor. In estimating biomass reaction rates, the level of biomass effective in

220

Table 7-3. Packed-Bed Denitrification (Partial Installation List)

| Facility and Location	Description of Packed-Bed Denitrification System	Capacity	
		L/s	mgd
Tampa, Florida Hookers Point AWTP	Twelve 97 m² (1,050 sq ft) filters Nineteen 93 m² (1,000 sq ft) filters	4,208	96.0
Seminole County, Florida NW Area Regional WW Facility Expansion	Two 46 m² (500 sq ft) filters	110	2.5
Port Orange, Florida	Six 52 m² (560 sq ft) filters	351	8.0
Hillsborough County, Florida Valrico Wastewater Facility	Three 46 m² (500 sq ft) filters	132	3.0
U.S. Home Brandon, Florida	One 19 m² (200 sq ft) filter	33	0.75
Purity Farms Clearwater, Florida	One 9.3 m² (100 sq ft) filter	10	0.23
Hillsborough County, Florida Dale Mabry AWTP	Five 60 m² (650 sq ft) filters	264	6.0
Piney Orchards, Maryland	Four 9.3 m² (100 sq ft) filters	53	1.2
Hillsborough County, Florida Falkenburg RD AWTP	Five 46 m² (500 sq ft) filters	264	6.0
Altamonte Springs, Florida	Seven 56 m² (600 sq ft) 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. Fiesta Village Fort Myers, Florida	Four 37 m² (400 sq ft) filters for nitrate reduction and SS removal	96 220	2.2 (Avg.) 5.0 (Peak)
Kanapaha Wastewater Treatment Plant Gainesville, Florida	Six 46 m² (500 sq ft) filters	308 770	7.0 (Avg.) 17.5 (Peak)
Parkland III Expansion Islip, New York	Deep-bed gravity denitrification-effluent polishing system including four 5.6 m² (60 sq ft) filters	11	0.26
Fairfield Village New York	Two 5.6 m² (60 sq ft) deep-bed sand filters for effluent polishing and denitrification	4	0.085
Southhampton Hospital Southhampton, New York	Two 4.7 m² (50 sq ft) deep-bed sand filters for effluent polishing and denitrification	4	0.1
Blue Ridge Condo. Medford Brookhaven, New York	One deep-bed sand filter system. System includes three deep-bed gravity filter cells 5.6 m² (60 sq ft) each	9	0.2
Parkland III Islip, New York	One deep-bed gravity filtration system for effluent polishing and denitrification. System includes four deep-bed filter cells 5.6 m² (60 sq ft) each	28	0.65
Parr Village Yaphank, New York	Three 4.8 m² (52 sq ft) deep-bed sand filters	20	0.45
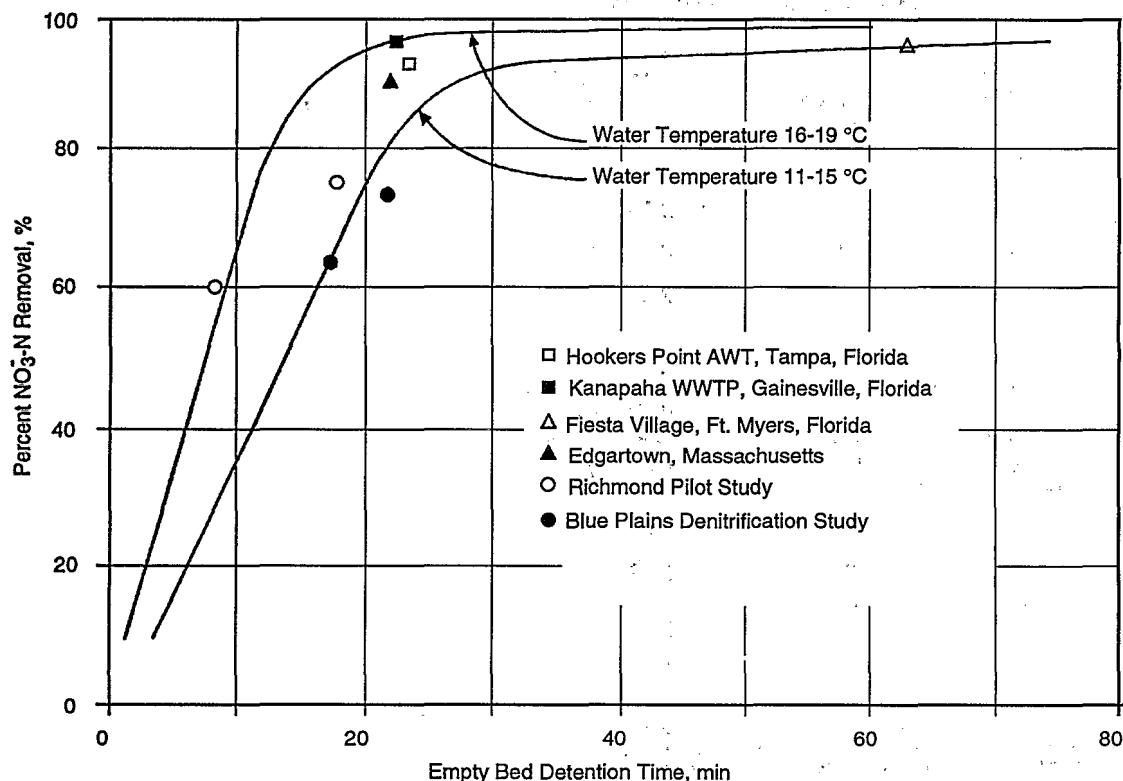
Figure 7-7. Typical design curves for empty-bed contact time (from Reference 14).

denitrification must also be known. One approach is to estimate the amount of biomass on the media surface and then use measured reaction rates per unit of biomass. This approach is of limited usefulness in design applications because there are insufficient data available at the present time to predict the level of biomass that will develop on the media. Biomass development is dependent on hydraulic regime, type of media, loading, means for promoting sloughing, and possibly the temperature of operation.

Denitrification rates can also be expressed in terms of nitrate removal rates per unit of filter surface. One approach uses mass of nitrate removed/unit area of filter surface/d (1). Table 7-4 shows rates for selected applications. (These rates are also given for each of the design examples presented in this section.) On this basis, high surface-removal rates would reflect extensive biological film development, whereas low surface-removal rates would reflect minimal surface film development. The surface denitrification rate varies considerably among the various denitrification column configurations and is affected by the loadings under which the process is operated.

As the applied nitrogen concentrations increase, the rate of nitrogen gas accumulation increases, forcing more frequent NRCs. As a practical matter, the maximum NRC frequency that can be tolerated is typically one every hour

(15). The relationship among nitrate concentration, loading, media underdrain configuration, and NRC is not well defined. The effect of influent nitrate concentrations and loading rate on effluent quality is shown in Figure 7-8 using data from actual operating experience at 20°C (68°F). As influent nitrate levels increase, the loading or application rate decreases for a given effluent criterion. Loading rates will also decrease predictably with decreased temperatures. Figure 7-9 illustrates the relationship between NO_3-N removal and loadings based on limited data from two installations. This curve was created by calculating the pounds of nitrate denitrified/sq ft/NRC (which ranged between 0.03 and 0.09) and then standardizing the concentration based on the concentration that could be denitrified between two NRCs, 2 hr apart (19). While the data presented in this curve are based on actual experience, recent media and underdrain modifications have decreased NRC frequency at some plants (8).

As discussed above, downflow packed-bed systems are effective in removing both SS and nitrogen. However, when considering the use of this process as both an effluent filter and a denitrification system, an important design factor—cost—should be kept in mind. It has been claimed that combining the functions of filtration and denitrification reduces tankage and equipment requirements and therefore yields cost savings in plants requiring filtration (19). However, column loading criteria are different

Table 7-4. Selected Downflow Packed-Bed Application Rates

Facility	Temp., °C	Media Depth, m	kg NOₓ-N removed/m² reactor/d	kg NOₓ-N removed/1,000 m² media surface area/d	kg NOₓ-N removed/1,000 m³ reactor/d[a]
Hookers Point[b]	20	1.7	2.7	1.6	1,600
Kanapaha[b]	NA	1.8	2.3	1.3	1,280
Dale Mabry[b]	24	1.8	0.54	0.29	290
Tetra Design Value[c]	15	1.8	2.4	1.4	1,330
Average			2.0	1.15	1,125

Facility	Temp., °C	Media Depth, ft	lb NOₓ-N removed/sq ft reactor/d	lb NOₓ-N removed/1,000 sq ft media surface area/d	lb NOₓ-N removed/1,000 cu ft reactor/d[a]
Hookers Point[b]	20	5.5	0.55	0.33	100
Kanapaha[b]	NA	6	0.48	0.27	80
Dale Mabry[b]	24	6	0.11	0.06	18
Tetra Design Value[c]	15	6	0.5	0.28	83
Average			0.41	0.23	70

[a] Based on 985 m²/m³ (300 sq ft of surface area/cu ft) of reactor.
[b] Based on operating data, design criteria may be higher.
[c] Based on design example in Chapter 7.
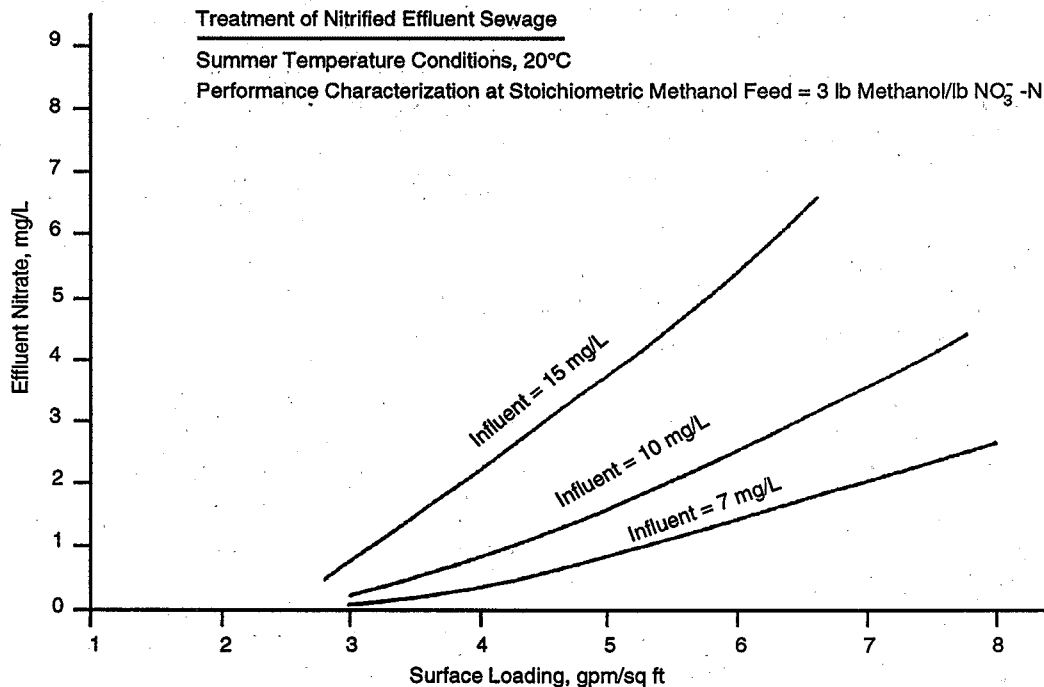
Figure 7-8. Effect of nitrate concentrations on loading rates in downflow packed-bed systems (from Reference 16).
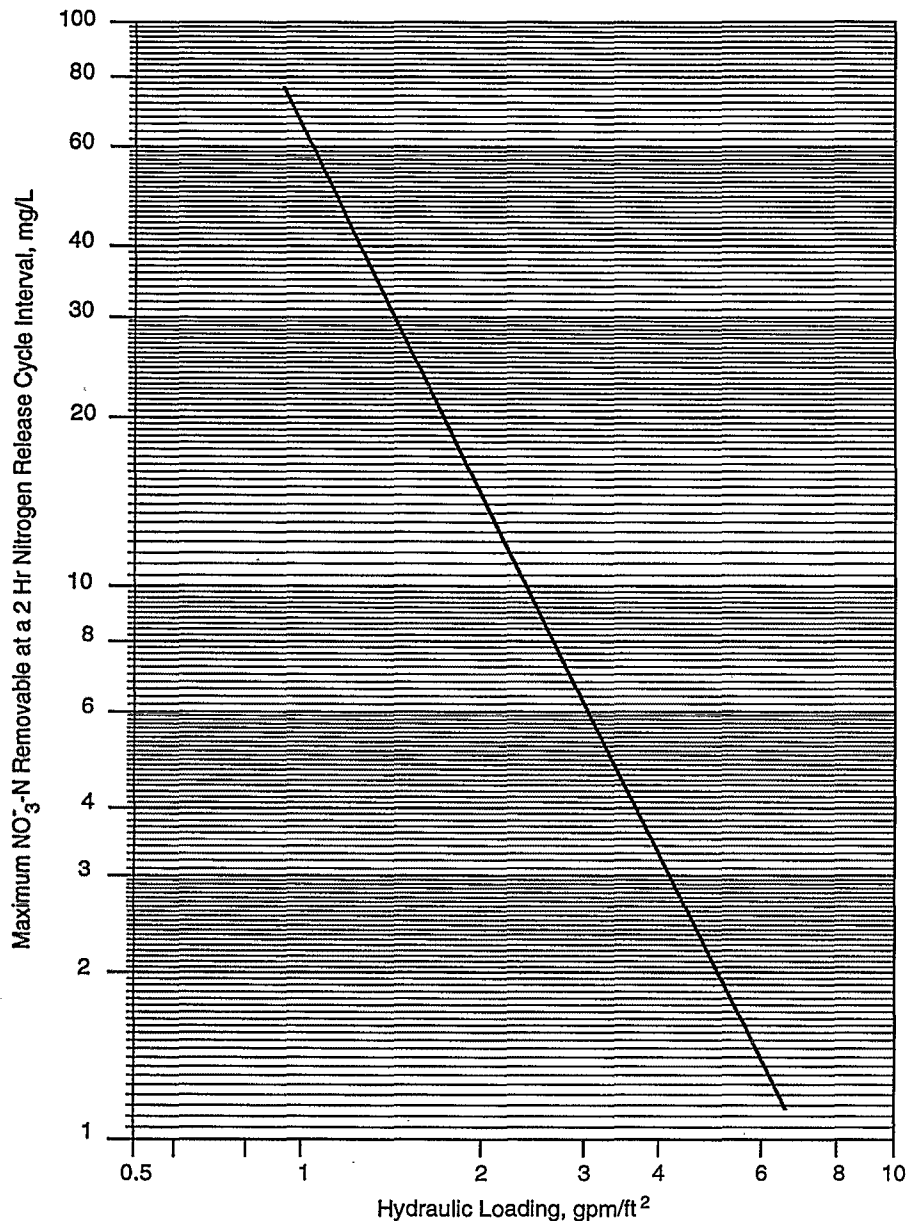
Figure 7-9. Downflow packed-bed denitrification performance (from References 17 and 18).

for the functions of filtration and nitrogen removal. For effluent filtration, fairly high hydraulic loadings can be applied (234–351 $m^3/m^2/d$ [4–6 gpm/sq ft]). However, for filters 0.9–1.8 m (3–6 ft) deep acting as denitrification columns, available data indicate that hydraulic loading should be between 29–88 $m^3/m^2/d$ (0.5–1.5 gpm/sq ft) at a wastewater temperature of 10°C (50°F) (1). Thus, to accomplish denitrification at 10°C (50°F), it could be necessary to have filter surface areas five times as large as those required for filtration alone. Consequently, an economic analysis must be done to determine the most economic process configuration for the particular system.

The physical design characteristics of downflow packed-bed systems are essentially the same as for any deep-bed filter system. Much like a typical filter, the downflow packed-bed filter system consists of three major components: the media, the underdrain, and the backwash facilities. Typical design criteria are shown in Table 7-5. No attempt has been made to discuss all factors in filter design. The reader is directed to the *Process Manual for Suspended Solids Removal* (20), to *Water Treatment Plant Design* (21), and to MOP 8 (8).

The media must be carefully selected to ensure proper size, shape, hardness, and density. Also, it must be large

Table 7-5. Typical Design Criteria for Downflow Packed-Bed System

Media Type	Sand: Effective size 1.8–2.3 mm Sphericity: 0.8–0.9 Specific gravity: 2.4–2.6 Dual media: coal also used
Media Depth	1.2–1.8 m (4–6 ft)
Backwashing	Duration: 5–10 min Water: 350–470 m/d (6–8 gpm/sq ft) Air: 1.5–1.8 m/min (5–6 cfm/sq ft)
Application Rates	60–120 m/d (1–5 gpm/sq ft)
Reseeding	Duration: 1-2 min Rate: 880–1,170 m/d (15–20 gpm/sq ft)
Nitrogen Release Cycle	Duration: Up to 5 min Interval: 1-6 hr Rate: 293 m/d (5 gpm/sq ft)
Flow Control	Variable declining rate or Influent flow control or Effluent flow control

enough to facilitate bed penetration but small enough to provide a high surface area for microorganisms and to prevent solids breakthrough (19). Media typically range in size from 1.8 to 2.3 mm. Coarser, deeper media filters are often used for denitrification rather than the final media filters that are typically found in municipal wastewater applications.

Rounded media are typically used because of the improved backwash characteristics of round grains, which tend to rotate during backwash and create a vigorous mutual scrubbing action. The scrubbing results in particularly clean media after backwashing. However, this agitation necessitates that the filter media be hard.

The density of the media is also a key factor. Bed expansion is a function of particle size, shape, density, and water flow rate. Higher density media require a higher flow rate to provide greater bed expansion. TETRA Technologies uses high-density sand with a specific gravity of 2.5 to minimize media loss during backwash because the denser media are more readily retained in the filter vessel.

While TETRA Technologies' underdrain system consists of precast concrete blocks, many other underdrain systems such as nozzles and high-density plastic distribution blocks (see Figure 7-10) have been used in filtering applications other than denitrification. These alternative underdrain systems are discussed more thoroughly elsewhere (22,23). The underdrains are designed to distribute the backwash water and air and to collect the filtered water. The backwash system must operate in conjunction with the underdrain, which distributes the backwash water evenly to the media. The backwash system consists of backwash water pumps and, if desired, backwash blowers. The manufacturer generally uses a combined air/water wash: for 15 min, air is pumped at 1.8 m^3/m^2/min (6 cfm/sq ft); water is pumped at 351 m^3/m^2/d (6 gpm/sq ft). The air/water wash is followed by a water-only wash at 468 m^3/m^2/d (8 gpm/sq ft), for 5 min. A separate air/water wash is also beneficial in scouring the media. Rates as high as 1,170 m^3/m^2/d (20 gpm/sq ft) would be required for full-bed expansion. Air entrainment in the influent and the media should be minimized so that nitrogen removal is not hindered. The water-only wash can be used to purge the media of entrained air. In a denitrification filter, this is particularly important because anoxic conditions must be maintained.
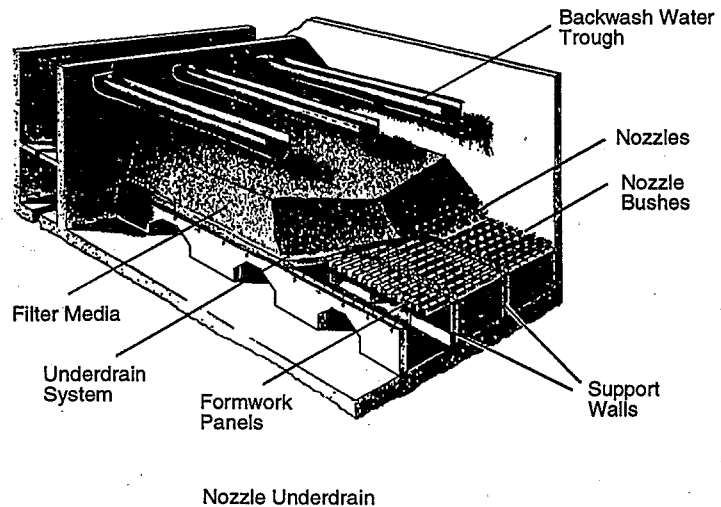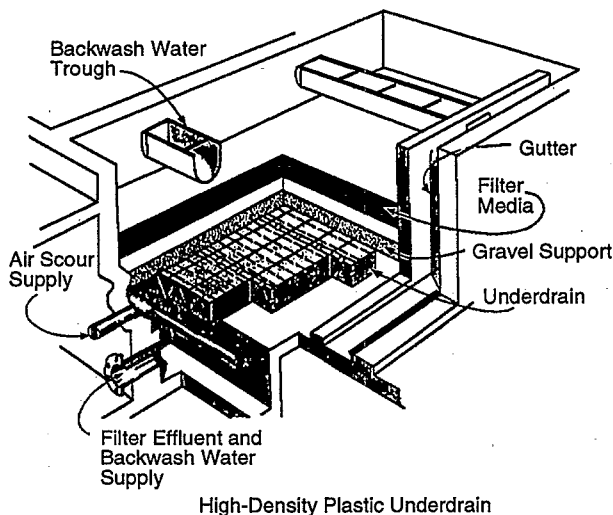

High-Density Plastic Underdrain

Nozzle Underdrain

Figure 7-10. Alternative filter underdrain systems.

Presenting useful design criteria based on kinetic reactions is difficult, partly because of the variability of the systems but also because of the proprietary nature of the process. However as explained below, there are some useful approaches for characterizing denitrification on attached growth systems.

7.3.2.1 Design Examples

The design examples that are presented for downflow packed-bed systems are based on the discussion in Chapter 4 about the kinetics of denitrification and the fundamental equations used in system design. These examples also draw from the influent criteria discussion in Chapter 2. The application rates used in this example are for illustration purposes only and are based on typical data; actual rates must be selected on a site-specific basis. The cases presented below involve different effluent requirements (see Table 7-6). A process flow schematic for a downflow packed-bed system is provided in Figure 7-11.

Table 7-6. Influent Characteristics and Design Effluent Limits for Denitrification Design Examples: Downflow Packed-Bed System

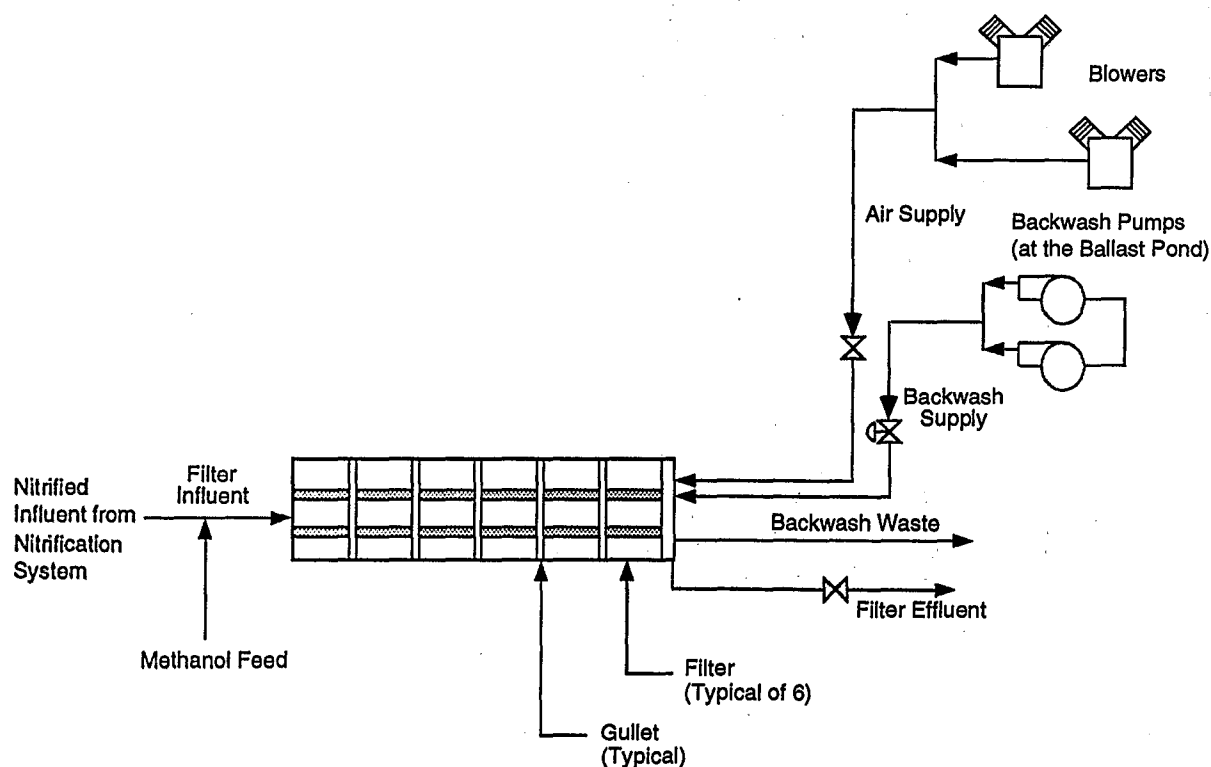| Characteristic	Nitrified Effluent	Case 1 Effluent Limits	Case 2 Effluent Limits
Minimum Monthly Temp.	—	15°C	15°C
Average Flow, m³/d	—	18,930	18,930
Average Flow, mgd	—	5	5
Peak Day Week, m³/d	—	28,400	28,400
Peak Day Week, mgd	—	7.5	7.5
TSS, mg/L	15	30	10
CBOD, mg/L	3	30	10
COD, mg/L	33	—	31
TKN, mg/L	1.8	—	—
$(NO_3^- + NO_2^-)$-N, mg/L	23.4	7	3
NH_4^+-N, mg/L	0.05	2	2
TN, mg/L	26.5	10	6

Figure 7-11. Design example schematic of downflow packed-bed system.

Design Example: Case 1—Downflow Packed-Bed Denitrification System

1. Calculate nitrate removed:

   (23.4 − 7 mg/L) (18,930 m³/d)/1,000 = 311 kg (684 lb) NO_x-N/d

   - Secondary effluent meets TSS discharge criteria, so TSS removal is not a parameter of concern; the filter will reduce the TSS to 5 mg/L

   - Calculate TSS removed:

     (15 − 5 mg/L) (18,930 m³/d)/1,000 = 189 kg (417 lb) TSS/d

2. Determine filter area based on nitrogen removal:

   - 2.45 kg NO_x-N/m²/d (0.5 lb NO_x-N/sq ft/d) is assumed based on TETRA Technologies operating experience at 15°C

     (311 kg NO_x-N/d) / (2.45 kg NO_x-N/m²/d) = 127 m² (1,368 sq ft) of filter

3. Calculation of filter area based on solids filtration is not needed for effluent limits.

   - Nitrogen removal conditions control

   - Assume: 127 m² (1,368 sq ft) of filter area

   - Determine number and size of filters

   - Assume: six filters, with five in service
     127 m²/5 filters = 25.4 m² (274 sq ft)/filter

   - Assume: six 3 m x 8.5 m (10 ft x 28 ft) filters with one out of service for backwash
     Total filter area = (5 filters) (25.5 m²/filter) = 127.5 m² (1,400 sq ft)

   - Calculate actual loading rate:
     (18,930 m³/d) ÷ (127.5 m²) = 148 m³/m²/d (2.52 gpm/sq ft)

4. Calculate backwash requirements:

   - Assume: 1) maximum backwash rate of 468 m³/m²/d (8 gpm/sq ft) on one filter; 2) average backwash rate of 351 m³/m²/min (6 gpm/sq ft) on one filter
     (25.5 m²) (468 m³/m²/d) / (1,440) = 8.3 m³/min (2,200 gpm) maximum backwash rate
     (25.5 m²) (351 m³/m²/d) / (1,440) = 6.2 m³/min (1,640 gpm) average backwash rate

   - Assume: backwash air rate of 1.8 m³/m²/min (6 cfm/sq ft) on one filter
     (25.5 m²) (1.8 m³/m²/d) = 45.9 m³/min (1,680 cfm) backwash air

   - Calculate volume of backwash flow:
     Duration = 15 min at average backwash flow and 5 min at maximum flow
     Volume = (6.2 m³/min)(15 min) + (8.3 m³/min)(5 min)
     = 134.5 m³ (35,400 gal)

5. Calculate backwash frequency:

   - Assume: 7.34 kg/m² (1.5 lb/sq ft) TSS removed/backwash (based on TETRA Technologies operating experience)
     Backwash = (189 kg TSS/d) ÷ [(7.34 kg TSS/m²) (127.5 m²/filter)] = 0.20 backwashes/d/filter
     —or—

   - One backwash every 120 hr/filter (with six filters, this is approximately one backwash per day)

   - The backwash volume generated per day is:
     (1 backwash/d) (134.5 m³/backwash) = 134.5 m³ (35,400 gal)/d

Design Example: Case 1 (continued)

This equals 0.7 percent of the average flow; therefore, the actual hydraulic application rate will be 0.7 percent higher (i.e., 148 m^3/m^2/d x 1.007 = 149 m^3/m^2/d).

6. Calculate frequency of nitrogen release cycle (NRC):

   - Assume: 0.245 kg NO_x-N/m^2/NRC (0.05 lb NO_x-N/sq ft/NRC) (Reference 18)

     (311 kg NO_x-N/d) / [(0.245 kg NO_x-N/m^2/NRC) (127.5 m^2)] = 9.9 NRC/d

     (24 hr/d) / (9.9 NRC/d) = 2.4 hr/NRC

     NRC frequency = 2.4 hr

7. Calculate methanol requirements.

   - Assume: 3 kg methanol/kg of NO_x-N removed.
     Methanol = (311 kg NO_x-N/d) (3 kg methanol/kg NO_x-N)
           = 933 kg (2,055 lb) methanol/d

     (933 kg/d) (1 L/0.79 kg) = 1,181 L (312 gal)/d

     Methanol dose = 49.2 L (13 gal)/hr

   - Alternatively, methanol requirements can be computed from Equation 4-10. If the influent DO to denitrification is assumed to be 3 mg/L, and minimal residual methanol, the methanol requirement is:

     (311 kg)(2.47) + (0.87)(3)(18,930 m^3/d/1,000) = 817 kg (1,800 lb)

8. Calculate biomass production

   - From suspended growth design example (Section 7.2.3.1), use 0.18 kg VSS/kg of COD removed

   - Assume: All COD removed is methanol

     COD removed = (933 kg methanol/d) (1.5 kg COD/kg methanol) = 1,400 kg/d
       VSS = (0.18 kg VSS/kg COD) (1,400 kg COD/d)
           = 252 kg (555 lb) VSS produced/d

   At 75 percent volatile:

   - TSS produced = 252 / 0.75 = 336 kg (740 lb) TSS/d

   - Compare to yield based on nitrate removed:
     Assume: Yield = 0.5 kg/kg of NO_x-N removed (TETRA Technologies recommendation).
     (311 kg NO_x-N/d) x (0.5 kg biomass/kg NO_x-N removed)
     = 156 kg (344 lb) biomass/d

9. Calculate approximate motor horsepower

   - Assume: Two backwash pumps are in service, with one on standby

   - Capacity of each = (8.3 m^3/min)/2 = 4.15 m^3/min = 1,100 gpm

   - Assume: Capacity is 4.5 m^3/min @ 9.14 m TDH (1,200 gpm @ 30 ft TDH)
           hp = [(Q)(TDH)(Sp. Gravity)] / [(3,960)(Pump Eff.)(Motor Eff.)]
     Where:
                 Q = 1,200 gpm
               TDH = 30 ft (assumed)
         Sp. Gravity = 1
         Pump EFF = 75% (assumed)
         Motor EFF = 90% (assumed)
               hp = [(1,200) (30) (1)] / [(3,960) (0.75) (0.90)]
                 = 13.5 hp
               Kw = (hp) (0.746) = 10.1 Kw

Design Example: Case 1 (continued)

- Calculate blower horsepower; assume that one blower is in service and one is on standby.
  Capacity = 45.9 m³/min (1,680 cfm)

  $$hp = [(w)(R)(T_1) \div (550)(n)(e)] [(P_2/P_1)^n - 1]$$

  Where:
  w = weight flow of air, lb/sec
  R = gas constant (53.5)
  T_1 = absolute inlet temperature, °R
  P_1 = absolute inlet pressure, psia
  P_2 = absolute outlet pressure, psia
  n = (k − 1)/k = 0.283 for air (k = 1.395 for air)
  e = efficiency (usual range for compressors is 70 to 80 percent; assume 70 percent)

  The density of air at 20°C (68°F) is 1.20 kg/m³ (0.075 lb/cu ft); therefore, the weight flow of air is:
  w = (45.9 m³/min) (1 min/60 sec) (1.20 kg/m³)
  = 0.92 kg (2.0 lb)/sec

- Assume an inlet temperature of 68°F, an inlet pressure of 14.7 psia, and an average outlet pressure of 22.7 psia (pressure may be 3 to 4 psi greater at startup).
  $$hp = [(2.0)(53.5)(460 + 86)/(550)(0.283)(0.7)] [(22.7/14.7)^{0.283} - 1]$$
  = 70.2 hp
  kW = (73.7) (0.746) = 52.3 kW

Design Example: Case 2—Downflow Packed-Bed Denitrification System

1. Calculate Nitrate Removed:

   (23.4 − 3 mg/L) (18,930 m³/d) ÷ (1,000) = 386 kg (850 lb) NO_x-N/d

- Calculate mass of TSS removed:

  Effluent requirement is 10 mg/L TSS; however, the filter will produce an effluent of approximately 5 mg/L TSS. See Case 1 for solids removal of 189 kg TSS/d (417 lb/d).

2. Determine filter area based on nitrogen removal:

   (386 kg NO_x-N/d) / (2.45 kg NO_x-N/m²/d) = 158 m² (1,700 sq ft) of filter

   As a conservative measure, an effluent nitrate concentration of 0 mg/L could also be assumed; removal would then be equal to 23.4 mg/L:

   (23.4 − 0 mg/L) (18,930 m³/d) ÷ (1,000) = 443 kg (976 lb) NO_x-N/d

   and the resulting filter area would be 181 m² (1,950 sq ft)

3. Calculation of filter area based on solids filtration is not necessary because of effluent limits. However, a typical average filtration rate would be 176 m³/m²/d (3 gpm/sq ft):

   (18,930 m³/d) / (176 m³/m²/d) = 108 m² (1,162 sq ft) of filter

   Therefore, nitrogen requirements control the filter sizing.

Determine number and size of filters; assume six filters, with five in service and one in backwash:

158 m^2/5 filters = 31.6 m^2 (340 sq ft)/filter
Assume six filters each 3 m x 10.7 m (10 ft x 35 ft) with one out of service for backwash.
   Total filter area = (5 filters) (3 m) (10.7 m) = 160.5 m^2 (1,750 sq ft)
   160.5 m^2/5 filters = 32.1 m^2 (345 sq ft) / filter
Calculate actual loading rate:
   (18,930 m^3/d) / (160.5 m^2) = 118 m^3/m^2/d (2.0 gpm/sq ft)

4. Calculate backwash requirements

   ● Assume: Maximum backwash rate of 468 m^3/m^2/d (8 gpm/sq ft) on one filter

   ● Assume: Backwash rate of 351 m^3/m^2/d (6 gpm/sq ft) on one filter
      (32.1 m^2) (468 m^3/m^2/d) / (1,440) =10.4 m^3/min (2,750 gpm) maximum backwash rate
      (32.1) (351) / (1,440) =7.8 m^3/min (2,070 gpm) average backwash rate

   ● Assume: Backwash air rate of 1.8 m^3/m^2/min (6 cfm/sq ft) on one filter
      (32.1 m^2) (1.8 m^3/m^2/d) = 57.8 m^3/min (2,100 cfm) backwash air

   ● Calculate volume of backwash water (see Case 1):
      Volume = (10.4 m^3/min) (5 min) + (7.8 m^3/min)(15 min)
           = 169 m^3 (44,600 gal)

5. Calculate backwash frequency (see Case 1):
   Backwash = (189 kg TSS/d) / [(7.34 kg TSS/m^2) (160.5 m^2)]
           = 0.16 backwashes/d/filter
            —or—
One backwash every 150 hr/filter or approximately one backwash/d.

6. Calculate frequency of nitrogen release cycle (NRC):

   ● Assume: 0.245 kg NO$_x$-N/m^2/NRC (0.05 lb NO$_x$-N/sq ft/NRC) (Reference 18)
      (386 kg NO$_x$-N/d) / [(0.245 kg NO$_x$-N/m^2/NRC) (160.5 m^2)] = 9.8 NRC/d
      (24 hr/d) / (9.8 NRC/d) = 2.5 hr/NRC
   NRC frequency = 2.5 hr

7. Calculate methanol requirements (see Case 1):

   ● Assume: 3 kg methanol/kg of NO$_x$-N removed
      Methanol = (386 kg NO$_x$-N/d) (3 kg methanol/kg NO$_x$-N)
              = 1,158 kg (2,550 lb) methanol/d

   ● Alternatively, use Equation 4-10 to compute methanol requirements.

8. Calculate biomass production (see Case 1):
   COD removed = (1,158 kg methanol/d) (1.5 kg COD/kg methanol) = 1,737 kg/d
      VSS = (0.18 kg VSS/kg COD) (1,737 kg COD/d)
          = 313 kg (688 lb) VSS produced/d
   At 75 percent volatile TSS:
   TSS produced = 313 / 0.75 = 417 kg (918 lb) TSS/d

9. Calculate motor horsepower (see Case 1 for methodology).

7.3.3 Case Studies

Two case studies of downflow packed-bed systems follow. Operating data for other downflow packed-bed systems are provided in Table 7-7.

7.3.3.1 Hookers Point Wastewater Treatment Plant (Tampa Florida)

The 4,208-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 schematic is shown in Figure 7-12.

The downflow packed-bed denitrification system consists of 20 filters measuring 3 m x 32 m (10 ft x 105 ft). Each filter is filled with 142 cm (56 in) of coarse sand (2.3 mm), loaded at an average rate of 59–117 $m^3/m^2/d$ (1–2 gpm/sq ft) and having an empty-bed contact time of 45 min at average flow.

The Hookers Point WWTP receives domestic wastewater, with a 30-percent contribution from breweries (13). 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). A summary of monthly plant operating data is provided in Table 7-8. The data show that an average month's effluent is below 3 mg/L TN 83 percent of the time, with an average over this three-year period of 2.33 mg/L. (Note that the effluent limit was changed to 3 mg/L TN in October 1990. Prior to that time, the limit was

4 mg/L TN in summer and 5 mg/L in winter.) The maximum month values appear to be attributable to decreased performance of the upstream nitrification system. 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/d (0.27 lb/sq ft/d). 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 percent and 99 percent, respectively.

7.3.3.2 Dale Mabry Advanced Wastewater Treatment Plant, Tampa, Florida

The flowstream of the 263-L/s (6-mgd) Dale Mabry Advanced WWTP consists of screening, grit removal, oxidation ditch with anoxic zone, secondary clarification, downflow packed-bed denitrification filters, post-aeration, and disinfection. A schematic is provided in Figure 7-13.

There are two oxidation ditches (HRT = 19.2 hr) and five secondary clarifiers. The solids are thickened on site and hauled off site by a contract hauler. The denitrification system consists of five filters, each 3 m (10 ft) wide by 19.8 m (65 ft) long, with 1.8 m (6 ft) of 2.3-mm sand. The design average daily filter loading is 123 $m^3/m^2/d$ (2.1 gpm/sq ft). The plant discharge standards are 5, 5, 3, 1 mg/L for BOD_5, TSS, TN, and TP, respectively, on an annual basis. The monthly limit for TN is 3.75 mg/L. The wastewater is primarily domestic, and the NO_x-N concentration in the inlet to the denitrifying filter is 6-15 mg/L.

A summary of monthly plant operating data is provided in Table 7-9. The data show that the Dale Mabry WWTP can consistently meet its effluent limit, with an average TN of 2.08 mg/L. Based on the available data, the plant's nitrate removal rate is 0.54 kg NO_x-N/m^2/d (0.11 lb/sq ft/d).

Table 7-7. Operating Data for Selected Downflow Packed-Bed Systems

Facility	Florida Location	Capacity L/s	Capacity mgd	Average Rate $m^3/m^2/d$	Average Rate gpm/ sq ft	Number of Denit. Filters	Filter Size	Media Depth m	Media Depth in	Media Size, mm
Hookers Point	Tampa	4,208	96.0	123	2.1	20	3 m x 32 m (10 ft x 105 ft)	1.47	54	2.3
Fiesta Village	Ft. Myers	220	5.0	117	2.0	4	3 m x 13.4 m (10 ft x 44 ft)	1.83	72	3.0
Altamonte Springs	Altamonte	548	12.5	123	2.1	7	3 m x 18.3 m (10 ft x 60 ft)	1.83	72	Dual Media
Faulkensand Road, Hillsborough Co.	Tampa	264	6.0	29	0.5	5	3 m x 15.2 m (10 ft x 50 ft)	1.22	48	3.0
Dale Mabry	Tampa	264	6.0	123	2.1	5	3 m x 19.8 m (10 ft x 65 ft)	1.83	72	2.3
Port Orange	Port Orange	526	12.0	123	2.1	7	3 m x 17.1 m (10 ft x 56 ft)	1.07	42	1.8

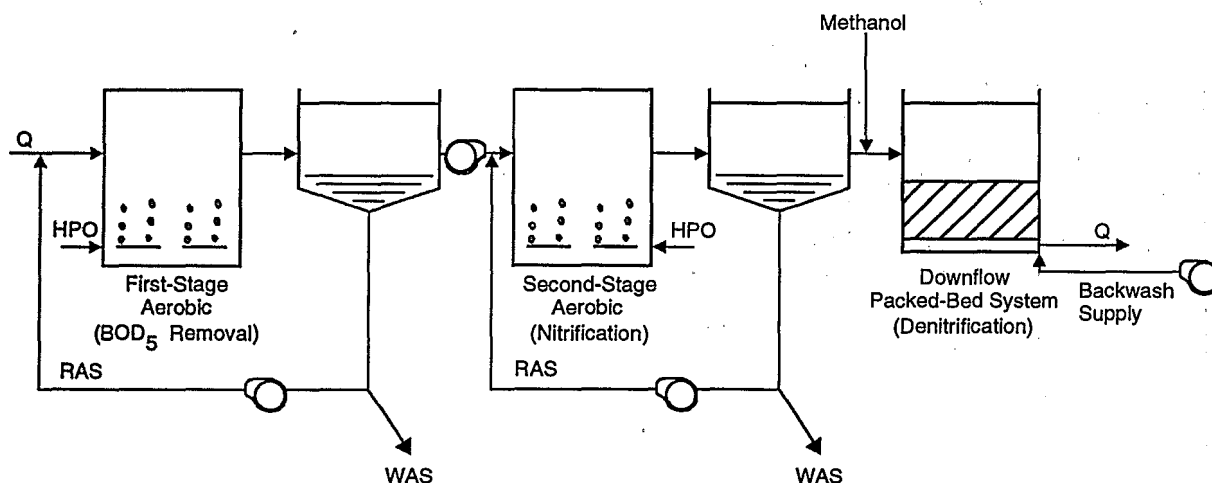Note: All plants have a 3 mg/L TN permit limit.

Figure 7-12. Schematic of Hookers Point Advanced Wastewater Treatment Plant.
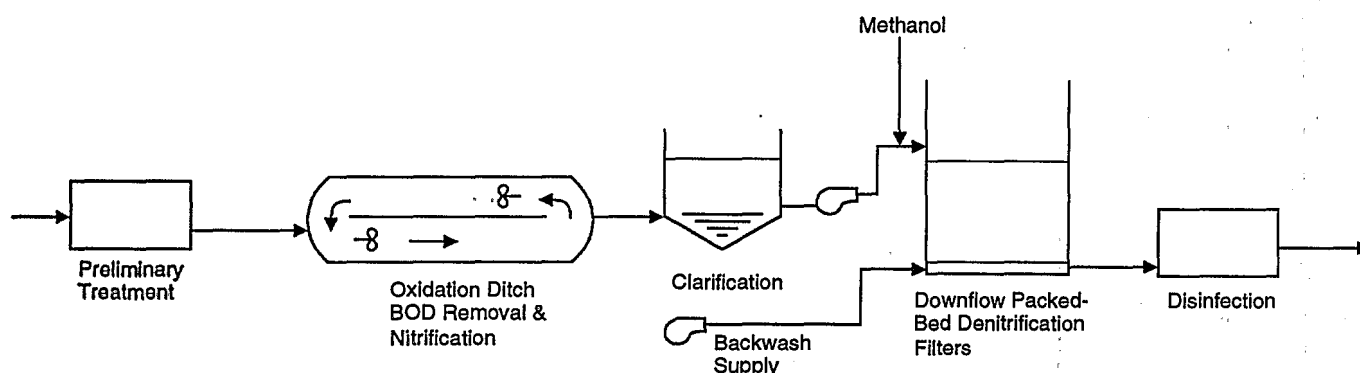

Figure 7-13. Schematic of Dale Mabry Wastewater Treatment Plant.

The nitrogen and SS removal efficiencies have been 94 percent and 99 percent, respectively, with effluent SS from 0.67 to 2.2 mg/L. These data are an indication of the favorable performance of the downflow packed-bed systems in regard to both SS and nitrogen removal. Variations in performance were generally attributable to difficulties with the methanol feed equipment; methanol feed pumps and a flow meter which paced the methanol feed were both replaced in 1991.

7.4 Upflow Fluidized-Bed Systems

7.4.1 Description

In a fluidized-bed reactor, nitrified secondary effluent passes upward through a column at a flow rate sufficient to produce a fluidized bed of media (typically sand). Denitrifying microorganisms attach to the sand, and the nitrified effluent passes through the media. Typically, small factory-assembled reactors are constructed as a column that is 2.4–4.3 m (8–14 ft) in diameter with varying height. The more common larger columns are fabricated in the

field from steel or concrete. Column heights are variable. Because the media are fluidized and the particles are not in contact with other particles, an extremely large surface-area-to-volume ratio is provided for supporting growth of denitrifying microorganisms. The main advantage of the fluidized-bed system is the small reactor size made possible by the increased efficiency associated with the development of a high biomass concentration, and the vertical configuration of the reactor. The specific surface area available for biological growth is from 244–305 m^2/m^3 (800–1,000 sq ft/cu ft) of reactor volume, as compared to approximately 91 m^2/m^3 (300 sq ft/cu ft) in a downflow packed-bed system. A typical flow schematic for a small, proprietary reactor is shown in Figure 7-14. Fluidized-bed reactors are discussed further in Chapter 6.

To date, fluidized-bed reactors have been installed at a limited number of facilities, where they have been implemented in aerobic, anoxic, and anaerobic modes. The systems have been used for a variety of applications such as treating chemical wastes, contaminated groundwater, fish hatchery wastes, municipal wastes, and high-strength industrial wastewater. Currently, the technology has been

Table 7-8. City of Tampa, Hookers Point Wastewater Treatment Plant: Monthly Performance Data

	Plant Influent					Denitrification Process Influent	Final Plant Effluent					
	Flow		BOD	TSS	TN	NO_x-N	BOD	TSS	TKN	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
JAN 89	2,362	53.9	383	220	32.3	16.8	2.0	0.9	1.47	0.00	0.13	2.80
FEB 89	2,266	51.7	299	238	32.3	17.5	3.0	1.9	1.81	0.03	0.91	2.72
MAR 89	2,209	50.4	275	214	34.5	18.8	2.0	1.6	1.75	0.22	1.72	3.47
APR 89	2,349	53.6	275	225	34.7	19.7	2.0	0.9	1.48	0.21	2.27	3.75
MAY 89	2,090	47.7	242	185	33.8	16.4	2.0	1.0	1.43	0.11	0.94	2.37
JUN 89	2,226	50.8	255	205	32.9	17.8	2.0	1.2	1.37	0.13	1.21	2.58
JUL 89	2,831	64.6	214	206	28.9	15.0	2.0	1.5	1.82	0.40	1.26	3.07
AUG 89	3,936	89.8	216	168	27	15.1	3.0	1.7	1.43	0.20	1.33	2.76
SEP 89	3,024	69.0	182	162	24.8	12.4	5.0	2.5	2.02	0.45	1.13	3.15
OCT 89	2,687	61.3	215	176	30	18.2	3.0	1.4	1.61	0.22	1.47	3.08
NOV 89	2,336	53.3	200	210	33.7	17.4	2.0	2.0	1.62	0.19	0.94	2.56
DEC 89	2,507	57.2	172	196	30.4	15.8	3.0	1.7	1.66	0.06	0.69	2.34
JAN 90	2,371	54.1	176	151	25.1	16.0	2.0	2.3	1.62	0.16	0.19	2.41
FEB 90	2,481	56.6	276	191	28.9	16.0	2.9	2.6	2.75	1.06	2.05	2.70
MAR 90	2,323	53.0	313	275	35.5	17.0	2.8	1.6	1.74	0.13	0.55	2.36
APR 90	2,612	60.0	280	236	36.2	18.0	2.8	1.2	1.62	0.12	0.60	1.22
MAY 90	2,108	48.1	241	273	37.2	17.0	3.1	1.5	1.59	0.14	1.37	2.85
JUN 90	2,261	51.6	268	278	34.1	16.8	2.0	1.5	1.41	0.16	1.33	2.83
JUL 90	2,709	61.8	231	167	27.8	16.8	1.3	2.8	0.08	0.30	0.31	2.22
AUG 90	2,577	58.9	222	171	29.7	8.5	2.2	2.1	2.45	0.94	1.61	3.06
SEP 90	2,319	52.9	256	191	38.6	11.9	2.0	1.8	2.76	1.38	1.36	3.32
OCT 90	2,297	52.5	267	183	32.3	13.9	2.9	1.8	1.40	0.02	1.46	1.86
NOV 90	2,980	68.0	222	201	36.1	14.4	3.8	2.9	1.58	0.03	0.41	1.99
DEC 90	1,937	44.2	281	197			3.8	0.9	1.65	0.48	0.48	2.12
JAN 91	2,314	52.8	247	184	32.1	13.8	2.0	1.9	1.49	0.02	0.44	1.86
FEB 91	2,104	48.0	227	281	34.9	14.4	3.0	2.0	1.58	0.03	1.41	1.39
MAR 91	2,814	64.2	281	293	35.5	13.4	3.0	1.8	1.89	0.04	0.50	2.62
APR 91	2,038	46.5	258	195	36.5	15.5	2.0	1.8	1.55	0.03	1.58	2.13
MAY 91	2,038	46.5	241	261	31.6	15.8	2.8	3.6	1.57	2.08	0.64	2.89
JUN 91	2,134	48.7	281	237	31.9	15.2	3.0	2.3	1.65	0.11	1.16	2.09
JUL 91	868	19.8	276	210	34.7	18.6	2.0	2.0	1.66	0.28	0.72	2.37
AUG 91	2,319	52.9	258	217	32.1	13.2	2.4	1.9	1.58	0.16	1.92	2.50
SEP 91	2,481	56.6	254	225	32.3	13.3	2.0	1.5	1.57	0.16	1.02	2.88
OCT 91	2,862	65.3	198	172	27.1	14.1	2.3	1.7	1.58	0.36	1.00	2.54
AVERAGE	2,253	51.4	253	218	34	14.42	2.64	2.0	1.68	0.34	1.01	2.33

Table 7-9. Dale Mabry Wastewater Treatment Plant: Monthly Performance Data

| Month	Plant Influent Flow L/s	Plant Influent Flow mgd	BOD mg/L	TSS mg/L	TN mg/L	Process Influent TSS mg/L	Process Influent NO$_x$-N mg/L	Final Plant Effluent BOD mg/L	Final Plant Effluent TSS mg/L	Final Plant Effluent NH$_4^+$-N mg/L	Final Plant Effluent NO$_x$-N mg/L	Final Plant Effluent TN mg/L
JAN 90	240	5.48	214	192	34.7	9	5.75	4.13	2.00	0.99	1.18	3.30
FEB 90	238	5.42	199	150	33.9	13	11.73	3.70	1.25	0.12	1.95	2.82
MAR 90	234	5.33	193	185	34.2	9	10.32	4.52	1.14	0.18	1.75	2.67
APR 90	231	5.27	204	189	35.2	10	9.11	2.02	2.23	0.15	1.79	2.34
MAY 90	230	5.24	195	169	34.5	8	6.70	2.22	2.61	0.23	0.97	1.99
JUN 90	238	5.42	182	142	32.2	5	6.82	2.85	2.20	0.47	1.30	2.58
JUL 90	241	5.50	182	152	32.2	9	11.30	2.20	1.74	0.11	1.57	2.46
AUG 90	239	5.46	183	156	32.3	3	11.20	2.11	1.68	0.10	1.30	2.03
SEP 90	236	5.39	179	166	32.1	6	10.10	2.26	1.43	0.11	0.61	1.12
OCT 90	234	5.33	185	154	33.1	2	11.03	1.78	1.00	0.12	0.97	1.61
NOV 90	232	5.29	210	182	35.7	3	10.95	1.75	0.81	0.12	1.34	2.12
DEC 90	235	5.37	218	180	34.6	5	9.07	1.91	1.77	0.23	1.04	1.94
JAN 91	222	5.06	188	165	34.4	3	8.25	2.02	0.67	0.16	1.38	2.73
FEB 91	213	4.87	188	146	33.7	4	8.52	2.05	1.04	1.16	1.02	2.88
MAR 91	155	3.54	220	188	34.7	4	12.25	1.36	1.02	0.22	1.43	2.27
APR 91	148	3.39	203	133	33.2	4	13.77	1.34	1.11	0.11	0.96	1.70
MAY 91	148	3.39	209	182	33.2	6	12.43	1.67	1.68	0.11	2.31	3.14
JUN 91	147	3.36	213	177	32.9	5	12.06	2.21	0.81	0.10	0.67	1.36
JUL 91	156	3.55	202	178	32.9	4	12.51	1.93	0.87	0.11	1.53	2.17
AUG 91	165	3.77	191	181	31.7	4	15.18	1.89	1.02	0.11	1.19	1.87
AVERAGE	191	4.36	201	169	33.5	4.2	11.34	1.85	1.10	0.22	1.20	2.08
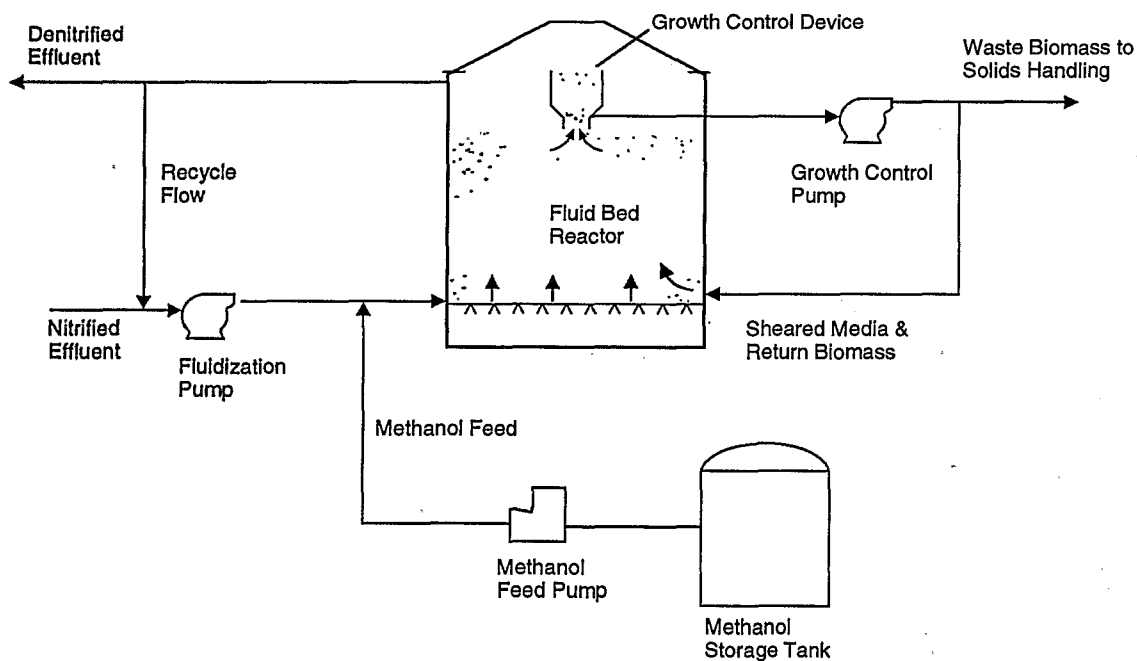
Figure 7-14. Schematic of upflow fluidized-bed system.

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 (24). However, as discussed below, there are successful municipal applications; Table 7-10 lists several industrial and municipal installations with fluidized-bed reactors. Some of these plants are not operating in the denitrification mode and others were designed solely for nitrification. Table 7-11 lists the loading at four of these plants, operating in the denitrification mode.

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 7-12 summarizes the types of reactors in use.

The principle of the fluidized-bed reactor is the same, regardless of the application. The uniformly round media become fluidized when the nitrified effluent passes upward through the bed. A thin active biomass forms on all sides of the media and causes the media to become less

Table 7-10. Full-Scale Applications of Upflow Fluidized-Bed Technology for Nitrogen Control[a]

Facility and Location	Application	Reactor Design Information (Approximate Dimensions)
Pensacola Wastewater Treatment Plant[b] Pensacola, FL	Denitrification of nitrified sanitary wastewater (not operating)	4 reactors, each 33.4 m² by 5.8 m high (360 sq ft by 19 ft)
Reno-Sparks Wastewater Treatment Plant[b] Reno, NV	Denitrification of nitrified sanitary wastewater	4 reactors, each 74.3 m² by 7.9 m high (800 sq ft by 26 ft)
Rancho California Wastewater Reclamation Plant[b] Riverside County, CA	Denitrification of nitrified sanitary wastewater (partially fluidized bed)	3 reactors, each 22.7 m² by 4.6 m high (244 sq ft by 15 ft)
IBM Plant[b] Somers, NY	Denitrification of nitrified sanitary wastewater	1 reactor, 2.8 m² by 4.3 m high (30 sq ft by 14 ft)
Anglian Water Authority Ipswich, England	Denitrification of river water to potable quality	1 reactor, 5.2 m² by 7.0 m high (56 sq ft by 23 ft)
Aquafuture Fish Hatchery Plant Turner Falls, MA	Ammonia removal from fish hatchery water	2 reactors, each 4.7 m² by 2.4 m high (51 sq ft by 8 ft)
Department of Energy Facility Fernald, Ohio	Denitrification of nuclear fuel processing wastewater	4 reactors, each 1.2 m² by 11.0 m high (13 sq ft by 36 ft)
Dworshak National Fish Hatchery Ahsahka, ID	Ammonia removal from fish hatchery water	7 reactors, each 14.3 m² (154 sq ft) by 7.3 m (24 ft) high (2 reactors) and 8.5 m (28 ft) high (5 reactors)
General Motors Plant Sandusky, OH	Aerobic carbonaceous oxidation of sanitary wastewater and aerobic carbonaceous oxidation and nitrification of wastewater from automotive manufacturing operations	3 reactors: two 8.8 m² by 7.0 m high (95 sq ft by 23 ft), and one 5.9 m² by 7.0 m high (64 sq ft by 23 ft)
General Motors Plant Ypsilanti, MI	Aerobic carbonaceous oxidation and nitrification of wastewater from automotive manufacturing operations	4 reactors, each 65.5 m² by 8.2 m high (705 sq ft by 27 ft)
Hazardous Waste Treatment Storage and Disposal Facility Hatfield, PA	Denitrification of industrial wastewater	2 reactors, each 10.5 m² by 7.6 m high (113 sq ft by 25 ft)
Sherwood Medical Plant Deland, FL	Denitrification of wastewater from cleaning operations during production of medical products	1 reactor, 3.2 m² by 5.5 m high (34 sq ft by 18 ft)

[a] Partial listing; includes only media-based fluidized-bed systems. Some plants are no longer operating and others are used for nitrification (25).
[b] Additional process design information is shown in Table 7-11.

235

Table 7-11. Process Design Information for Upflow Fluidized-Bed Systems (from Reference 24)

Parameter[a]	Facility			
	Pensacola[b,c]	Reno-Sparks	Rancho, CA[c]	IBM[c]
Mean wastewater flow, L/s	1,052	1,883	263	113
Mean wastewater flow, mgd	24	43	6[d]	1[d]
Maximum wastewater flow, L/s	1,490	2,400		
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] $m^3/m^2/d$	672	550	336	578
Hydraulic loading rate,[f] gpd/sq ft	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] See Table 7-10 for additional information.
[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 7-12. Types of Fluidized-Bed Denitrification Systems (from Reference 26)

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

dense than clean media. The lower density particles (i.e., the particles with the thickest biomass film) move toward the top. The open nature of the fluidized bed minimizes the chance for bed plugging.

7.4.2 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/d (400 lb/1,000 cu ft/d) (27). Other loading rates are shown in Table 7-13.

When the fluidized bed system is operated for denitrification, methanol is fed to the nitrified influent by injection into the recycle line (see Figure 7-14). 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. The amount of settled media depth is a func-

Table 7-13. Selected Upflow Fluidized-Bed Loading Rates

Facility and Location	Temp., °C	Reactor Height, m	kg NO_x-N removed/1,000 m^2 media surface area[a]/d	kg NO_x-N removed/1,000 m^3 reactor/d	Reference
Mineola, NY[b]	20	4.7	2.0	5,380	28
Reno-Sparks, NV[b]	13	6.2	0.78	2,010	
Rancho, CA[b]	25	1.2	1.17	3,000	29
Envirex[c]	15	2.2	3.4	6,420	
Average			1.84	4,200	

Facility and Location	Temp., °C	Reactor Height, ft	lb NO_x-N removed/1,000 sq ft media surface area[e]/d	lb NO_x-N removed/1,000 cu ft reactor/d	Reference
Mineola, NY[b]	20	15.5	0.41	335	28
Reno-Sparks, NV[b]	13	20.5	0.16	125	
Rancho, CA[b]	25	4	0.24	187	29
Envirex[c]	15	7.5	0.7	400	
Average			0.38	262	

[a] Based on 2,625 m^2/m^3 (800 sq ft of surface area/cu ft) of reactor.
[b] Based on actual plant data; design criteria may be different.
[c] Based on published Envirex literature; see design example in this chapter.

tion of the type and strength of waste being treated and is variable between 30 and 60 percent of bed volume, with a typical value of 50 percent.

Design and operating parameters that are important for maintaining high biomass concentrations in the reactor include (29): 1) type, size, and volume of media, 2) media density, 3) reactor cross sectional area, 4) biofilm thickness, and 5) expanded bed height, including expansion due to hydraulic flow and biofilm growth. The first three items can be established during design; however, items 4 and 5 become important operating parameters that affect fluidization and biomass concentrations.

The typical media sand is hard and round, with a size range of 0.3–0.6 mm. Its uniformity coefficient is 1.25–1.50. To ensure proper fluidization at the given flow rates, the particle density is important. Sands with a specific gravity of 2.6 have been used.

Two interrelated factors are the control of biofilm thickness and expanded bed height. An important design consideration is proper bed height selection and the media-biomass growth control system. The fluidized-bed system results in biofilm growth; the effective diameter of the particles increases and the density is reduced. As a result, the bed expands beyond the expansion of clean media. The lower density, larger particles (i.e., particles with the thickest biomass film) move toward the top. Controlling the thickness of the film is necessary to prevent media from being transported out of the reactor. A mechanical film thickness or growth-control device is used at the top of the reactor to, in effect, control bed expan-

sion. When the bed height reaches the bottom of the growth-control device, the growth-control pump is designed to start automatically. The growth-control system must minimize media carryover from the reactor to the effluent stream, minimize media loss and abrasion or contribution to SS in the effluent or recycling streams, and ensure that the returned media are clean.

The growth control system removes the upper layers of the media from the top of the reactor. The slush, or sheared biomass, and media slurry are then passed through a separation device, and the media are returned to the reactor. The sheared media have a thinner layer of biomass and are returned to the reactor to retain the proper amount of microorganisms. The sheared biomass is wasted to the solids-handling system. The clear zone above the expanded media is usually the height of the growth-control mechanism plus a freeboard of 2.4–3 m (8–10 ft). The clean media returned to the reactor have an effective biofilm thickness of less than 20 mm (30). The biomass wasted from the Envirex growth-control system is typically 0.5–1.5 percent solids. All known full-sized systems use a pump to shear the biomass from the media. The downstream mechanisms that separate the media-biomass slurry include (26):

- External wedge bars
- Screens with a spray wash
- Hydraulic separators internal to the reactor
- External cyclones
- Patented separation system (Ecolotrol/Envirex)

• Combination of the above

The most significant problems concerning the above systems are bed media loss and the release of biomass into the effluent (24). In other words, the systems may not adequately separate the sheared biomass and the media. As a result, either the media remain in the waste biomass and are lost, or the sheared biomass being returned with the media is discharged with the effluent. The ideal media-separation system must ensure there is no media carry-over from the reactor into the effluent and recycle stream. The system must not cause loss of media or contribute biomass SS to the effluent and recycle stream.

Excess biomass is also removed from the media by continuous sloughing off through death and abrasion. Sloughed off biomass can be a problem, but in a properly designed system, the effluent SS will generally be 15–20 mg/L for municipal wastewater applications.

The influent distribution manifold is another critical design feature of the fluidized-bed system. The manifold must:

• Achieve uniform distribution of flow across the entire reactor area to maximize reaction kinetics.

• Prevent plugging and media escape.

• Minimize abrasive wear.

• Maintain uniform flow throughout the media depth and minimize the movement of biomass through the reactor.

• Minimize shearing of biomass above the influent distribution manifold, thereby promoting uniform biofilm buildup throughout the media.

Most problems with distribution manifolds can be attributed to plugging. However, plugging can be prevented by removing solids from the influent stream and with a hydraulic design that prevents media backflow into the distribution manifold (24). The most common influent distribution systems consist of a header manifold pipe with lateral pipes branching out. Nozzles attached to the laterals point downward at or near the reactor bottom.

A key advantage of the fluidized-bed system is its mobility and small footprint, which is conducive to factory assembly of small reactors. Therefore, some reactors are assembled in the factory and have no moving parts; the methanol feed system, growth-control pump, and influent pump are the only required mechanical components. However, the majority of installed systems are in-ground concrete units or above-ground steel units that are fabricated on site. The main disadvantages of the fluidized-bed system are the limitations on reactor size, energy requirements (dominated by high recycle rates), difficulties in biomass control and media selection (media loss and biomass in the effluent), and the imprecision in process control because of difficulty in monitoring biomass concentration.

The design of fluidized-bed reactors is based on a combination of hydraulic and biological kinetic models. A number of models have been suggested (31–33). A half-order kinetic model proposed by Stephenson and Murphy (33), combined with an Arrhenius temperature relationship (1), was reported to adequately describe denitrification of municipal wastewater over a range of temperatures. This results in a curve similar to that shown in Figure 7-15.

7.4.3 Design Examples

The design approaches presented here for upflow fluidized-bed systems are based on the discussions in Chapter 4 about the kinetics of denitrification and the fundamental equations used in system design. As with preceding system design examples in this chapter, these draw from the theory discussed in Chapter 4 and the influent criteria discussed in Chapter 2. The application rates used are for illustration purposes only and are based on recommendations from Envirex. The use of more conservative rates may be appropriate, depending on the system application. Onsite pilot testing is recommended to generate design criteria (see Chapter 6). The examples use two different effluent requirements. The calculations for Case 1 are presented in more detail than those for Case 2; thus Case 1 should be referred to for a complete methodology. Table 7-14 lists the nitrified effluent characteristics (influent to the denitrification reactor), the final effluent limits, and the values used for design. A process flow schematic for an upflow fluidized-bed system is provided in Figure 7-16.

Table 7-14. Upflow Fluidized-Bed System Design Examples—Influent Characteristics and Effluent Limits

Characteristic	Nitrified Effluent	Case 1 Effluent Limits	Case 2 Effluent Limits
Minimum Monthly Temp.	—	15°C	15°C
Average Flow, m³/d	—	18,930	18,930
Average Flow, mgd	—	5	5
Peak Week Flow, m³/d	—	28,396	28,396
Peak Week Flow, mgd	—	7.5	7.5
TSS, mg/L	15	30	10
CBOD, mg/L	3	30	10
COD, mg/L	33	—	31
TKN, mg/L	1.8	—	—
$(NO_3^- + NO_2^-)$-N, mg/L	23.4	7	3
NH_4^+-N, mg/L	.05	2	1
TN, mg/L	26.5	10	5

238

Design Example: Case 1–Upflow Fluidized-Bed Denitrification System

1. Calculate nitrate removed:

   (23.4 − 7 mg/L) (18,930 m³/d) / 1,000 = 311 kg (684 lb) NO_x-N/d

2. Calculate reactor volume:

   - Assume:

     Loading rate = 6.42 kg NO_x-N/m³/d (400 lb NO_x-N/1,000 cu ft/d) [See Figure 7-15.]

   - Calculate volume of reactor:

     Volume = (311 kg NO_x-N/d) / (6.42 kg NO_x-N/m³/d)
              = 48.4 m³ (1,710 cu ft)

   - Assume: 3.65-m (12-ft) diameter reactor

     Area = π (3.65)²/4 = 10.5 m² (113 sq ft)/reactor

   - Assume: Two reactors in service and one on standby

   - Calculate bed height:

     (48.4 m³) / [(10.5 m²/reactor) (2 reactors)] = 2.3 m (7.6 ft)

   - Use 3 m (10 ft) high bed with 1.8 m (6 ft) of freeboard for solids separation for a 4.9 m (16 ft) high reactor, based on manufacturer's standard.

     Total volume of in-service reactors = 2(10.5 m²) (3 m)

     = 63 m³
     = 2,225 cu ft
     = 16,640 gal

3. Calculate HRT at average flow:

   (63 m³)(1,440 min/d)/18,930 m³/d = 4.8 min

   Calculate HRT at peak flow:

   (63 m³)(1,440 min/d)/28,396 m³/d = 3.2 min

4. Check flux rate:

   Flux rate = flow/area

   (18,930 m³/d) / (10.5 m²) / (2) = 901 m³/m²/d (15.3 gpm/sq ft) at average flow

   (28,396 m³/d) / (10.5 m²) / (2) = 1,352 m³/m²/d (23.0 gpm/sq ft) at peak week flow

5. Calculate actual nitrogen loading based on selected reactor:

   (311 kg NO_x-N/d) / (63 m³) = 4.94 kg NO_x-N/m³/d (308 lb NO_x-N/1,000 cu ft/d)

6. Calculate recycle rate:

   - Maintain reactor flow rate equal to peak flow rate of 19.72 m³/min (5,208 gpm) since the peak flow rate provides adequate fluidization for the media in this example. Flux rate should be between 880–1,470 m³/m²/d (15–25 gpm/sq ft).

   - Recycle required at average flow is equal to 19.72 −13.15 = 6.57 m³/min (1,736 gpm).

7. Calculate methanol required:

   - Calculated nitrate removed = 311 kg/d (684 lb/d)

   - Assume: 3 kg methanol/kg of NO_x-N removed

     Methanol = (311 kg NO_x-N/d) (3 kg methanol/kg NO_x-N)
              = 933 kg (2,055 lb) methanol/d

239

Design Example: Case 1 (continued)

(933 kg/d) (1 L/0.79 kg) = 1,181 L (312 gal)/d

Methanol dose = 49.2 L (13 gal)/hr

- Alternatively, methanol requirements can be computed from Equation 4-10. If the influent DO to denitrification is assumed to be 3 mg/L, the methanol requirement (neglecting residual methanol) is:

(311 kg)(2.47) + (0.87)(3)(18,930 m^3/d) / 1,000 = 818 kg (1,800 lb)

8. Calculate biomass produced:

- From the suspended growth design example (Section 7.2.3.1), use 0.18 kg VSS/kg COD removed. Assume that all COD removed is methanol:

COD removed = (933 kg methanol/d) (1.5 kg COD/kg methanol) = 1,400 kg/d
VSS = (0.18 kg VSS/kg COD) (1,400 kg COD/d)
= 252 kg (485 lb) VSS produced/d

At 75 percent volatile:

TSS produced = 252 / 0.75 = 336 kg (741 lb) TSS/d = 1.08 kg TSS/kg NO$_3^-$-N

- Typical biomass production is 0.4–0.8 kg TSS/kg NO$_x$-N removed

- Calculate excess biomass flow rate (assume 1 percent solids):
Flow = (336 kg TSS/d) (1/0.01) (1 L/ kg) (1/1,440 min/d)
= 23.3 L/min (6.2 gpm)

9. Calculate horsepower of pumps:

- Fluidization pump:
Total pump capacity is 19,720 L/min (5,208 gpm) (peak week flow)
With 2 pumps: 19,720/2 = 9,860 L/min (2,605 gpm)

- A typical fluidized-bed configuration requires a fluidization pump with approximately 12.2 m (40 ft) TDH. This should be verified for the actual reactor configuration.
hp = [(Q)(TDH)(Sp. Gravity)] / [(3,960)(Pump Eff.)(Motor Eff.)]
Where:
TDH = 40 ft (assumed)
Pump EFF = 75% (assumed)
hp = [(2,605) (40) (1)] / [(3,960) (0.75) (0.90)]
= 39.0 hp
Kw = (hp) (0.746) = 29.0 Kw

- Calculate horsepower of growth control pump:
Base flow rate on biomass flow rate of 3 gpm for each pump.
hp = [(3) (40) (1)] / [(3,960) (0.75) (0.90)]
= 0.04 hp
Kw = (hp) (0.746) = 0.03 Kw

Design Example: Case 2—Upflow Fluidized-Bed Denitrification System

1. Calculate nitrate removed:

   $(23.4 - 3$ mg/L$)$ $(18,930$ m^3/d$)$ $\div$ 1,000 = 386 kg (850 lb) NO_x-N/d

2. Calculate reactor volume:

   * Assume:

     Loading rate = 6.42 kg NO_x-N/m^3/d (400 lb NO_x-N/1,000 cu ft/d) [See Figure 7-15.]

   * Calculate volume of reactor:
     Volume = (386 kg NO_x-N/d) / (6.42 kg NO_x-N/m^3/d)
     $\qquad$ = 60.1 m^3 (2,120 cu ft)

   * Assume: 3.65-m (12-ft) diameter reactor

     Area = π $(3.65)^2$/4 = 10.5 m^2 (113 sq ft)/reactor

   * Assume: Three reactors in service and one on standby

   * Calculate bed height:

     (60.1 m^3) / [(10.5 m^2/reactor) (3 reactors)] = 1.9 m (6.2 ft)

   * Use 3 m (10 ft) high bed with 1.8 m (6 ft) of freeboard for solids separation for a 4.9 m (16 ft) high reactor, based on manufacturer's standard.

     Total volume of in-service reactors = 3(10.5 m^2)(3 m)
     $\qquad\qquad\qquad\qquad\qquad$ = 94.5 m^3
     $\qquad\qquad\qquad\qquad\qquad$ = 3,340 cu ft
     $\qquad\qquad\qquad\qquad\qquad$ = 24,970 gal

3. Calculate HRT at average flow:

   (94.5 m^3)(1,440 min/d)/18,930 m^3/d = 7.2 min
   Calculate HRT at peak flow:
   (94.5 m^3)(1,440 min/d)/28,396 m^3/d = 4.8 min

4. Check flux rate at average flow:

   Flux rate = flow/area
   18,930 m^3/d / [(10.5 m^2) (3)] = 601 m^3/m^2/d (10.2 gpm/sq ft) at average flow
   28,396 m^3/d / [(10.5 m^2) (3)] = 901 m^3/m^2/d (15.3 gpm/sq ft) at peak flow

5. Calculate actual nitrogen loading based on selected reactor:

   (386 kg NO_x-N/d) / (94.5 m^3) = 4.1 kg NO_x-N/m^3/d (250 lb NO_x-N/1,000 cu ft/d)

6. Check recycle rate (see Case 1).

7. Calculate methanol required (see Case 1).

8. Calculate biomass production (see Case 1).

9. Calculate horsepower of pumps (see Case 1).

The process flow schematic is similar to Figure 7-16, except that for this example four reactors are used.
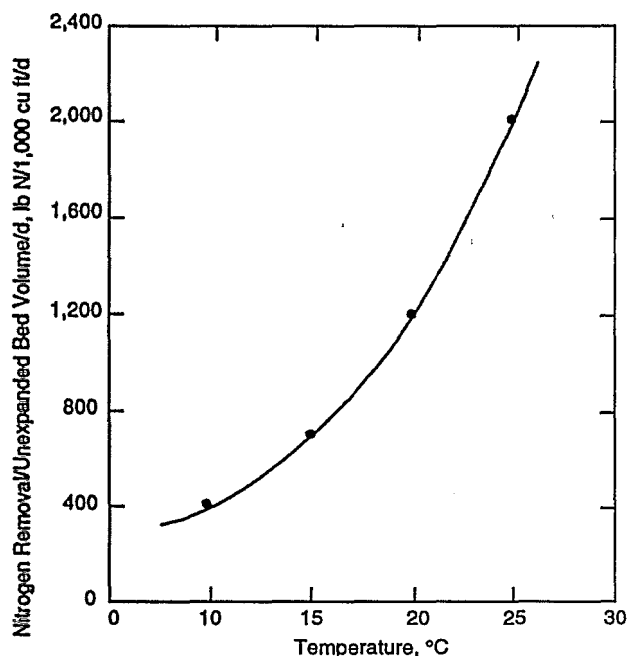
Figure 7-15. Temperature vs. loading rate for upflow fluidized-bed system (from Reference 1).

7.4.4 Case Studies

7.4.4.1 Reno-Sparks Wastewater Treatment Facility, Cities of Reno and Sparks, Nevada

A flow schematic for the 1,753-L/s (40-mgd) Reno-Sparks Wastewater Treatment Facility is shown in Figure 7-17. The plant consists of preliminary treatment, primary treat-ment, phosphorus and BOD removal in a sidestream phosphorous-removal system, nitrification biotowers, de-nitrification upflow fluidized-bed reactors, post aeration, effluent filtration, and disinfection. The solids-handling system consists of thickening, anaerobic digestion, and dewatering.

The denitrification system consists of four upflow fluidized-bed towers measuring approximately 8.2 m (27 ft) in di-ameter 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 (θ_c) is 8.5 d. The denitrification system, manufactured by Envirex, was designed to produce efflu-ent with a nitrate level of 2 mg/L. A summary of monthly plant operating data is provided in Table 7-15. The data indicate that the Reno-Sparks plant has consistently met its effluent requirements, with an average effluent ammo-nia 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 percent. The removal rate of the fluidized-bed reactors has been 6.4 kg NO_x-N/m^2/d (1.3 lb/sq ft/d), 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. Note that the design loadings are lower than those used in the design example.

7.5 Methanol Handling, Storage, Feed Control, and Excess Methanol Removal

Methanol is a chemical not normally dealt with in the operation of wastewater treatment plants. Care must be exercised in the design and operation of methanol han-dling, storage, and feeding facilities to ensure the safe and proper use of the methanol.
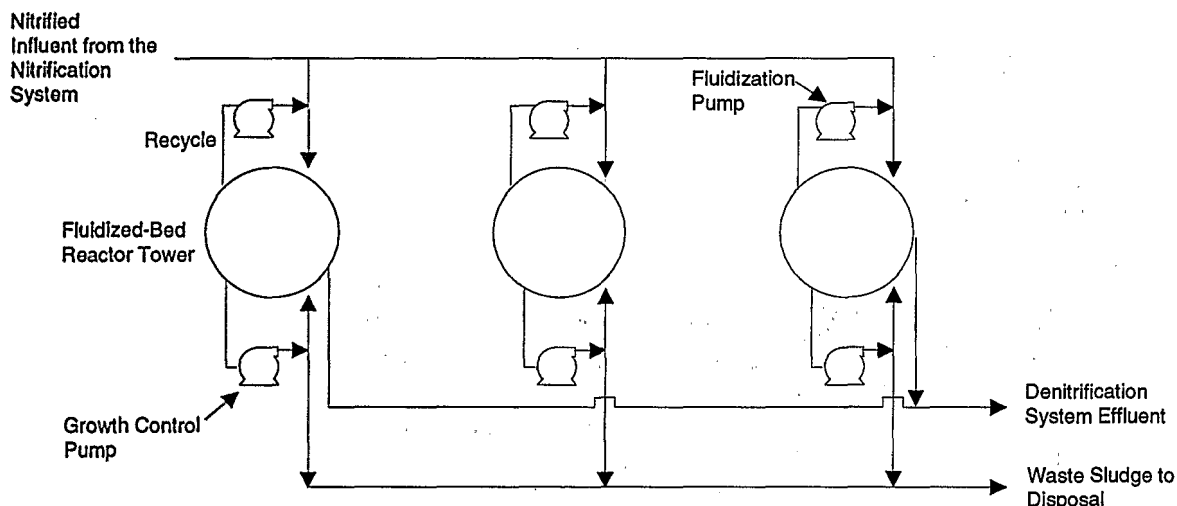

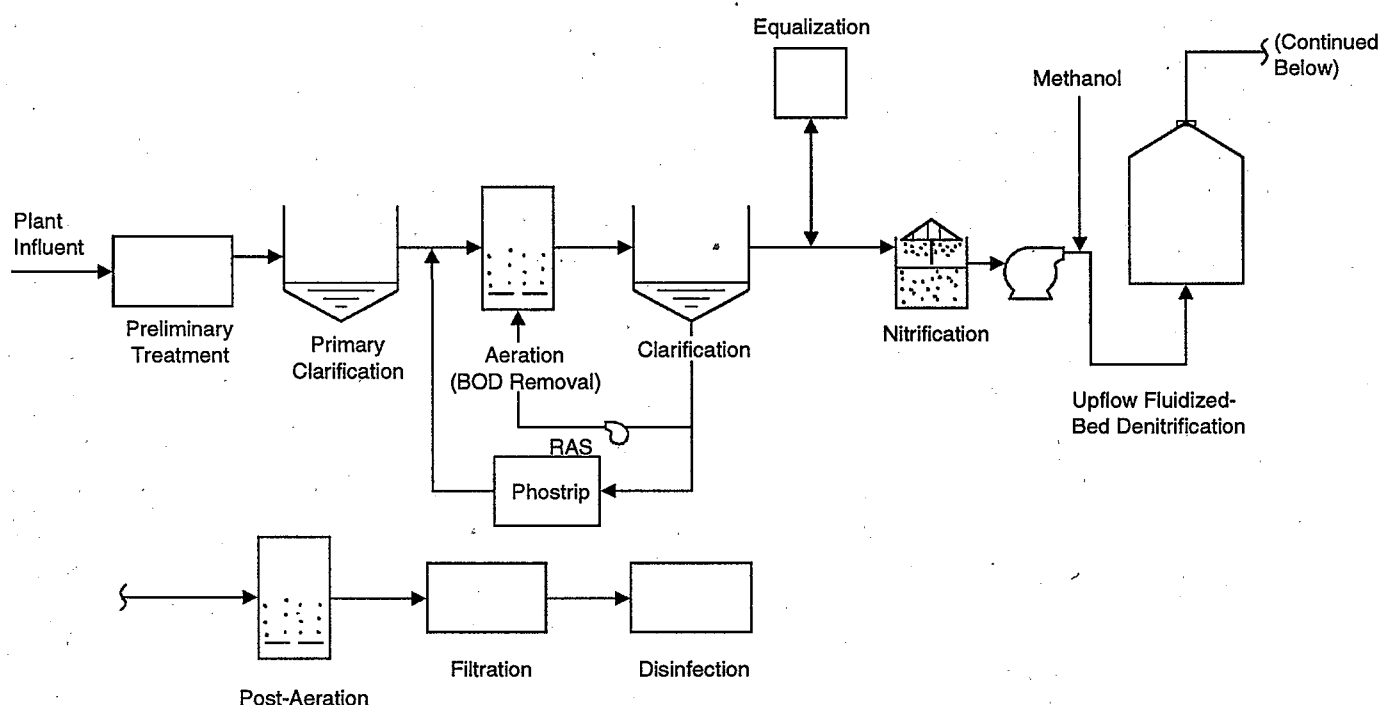Figure 7-16. Design example schematic of upflow fluidized-bed system.

Figure 7-17. Schematic of Reno-Sparks Wastewater Treatment Plant.

Table 7-15. Reno-Sparks Wastewater Treatment Plant: Monthly Performance Data

Month	Plant Influent Flow L/s	Plant Influent Flow mgd	BOD₅ mg/L	NH₄⁺-N mg/L	TN mg/L	Denitrification Process Influent NH₄⁺-N mg/L	Denitrification Process Influent NO_x-N mg/L	Denitrification Process Effluent TSS mg/L	Denitrification Process Effluent NO_x-N mg/L	Final Plant Effluent BOD mg/L	Final Plant Effluent TSS mg/L	Final Plant Effluent NH₄⁺-N mg/L	Final Plant Effluent NO_x-N mg/L	Final Plant Effluent TN mg/L
JAN 90	1,133	25.86	167			0.37	14.44		2.12	5	5	0.13	0.79	2.53
FEB 90	1,153	26.30	171			0.53	14.58		3.91	8	5	0.24	0.43	2.25
MAR 90	1,174	26.79	178			0.31	14.80		0.80	5	4	0.27	0.88	2.79
APR 90	1,156	26.37	177			1.37	14.21		0.82	6	2	0.29	0.90	3.55
MAY 90	1,169	26.66	183			0.67	12.84		0.19	5	3	0.08	0.15	1.38
JUN 90	1,221	27.87	175			0.35	14.02		0.06	3	2	0.11	0.16	1.43
JUL 90	1,202	27.42	152			0.30	14.07		0.04	3	2	0.10	0.06	1.11
AUG 90	1,233	28.14	164			0.91	13.70		0.29	6	10	0.32	0.14	1.82
SEP 90	1,211	27.64	168			0.56	14.18		0.12	7	18	0.41	0.11	2.50
OCT 90	1,175	26.81	149			0.15	11.03		0.03	3	2	0.05	0.01	1.01
NOV 90	1,144	26.10	162			0.37	13.17		0.10	4	3	0.19	0.07	1.26
DEC 90	1,192	27.20	150			1.13	13.92		0.22	4	6	0.33	0.37	2.00
JAN 91	1,168	26.64	132	21.1	32.3	0.45	14.04	22	0.24	3	4	0.17	0.35	1.64
FEB 91	1,122	25.60	130	22.2	32.3	0.46	14.26	20	0.16	3	3	0.18	0.28	1.58
MAR 91	1,149	26.21	135	22.5	32.2	0.35	14.88	18	0.39	3	4	0.10	0.42	1.70
APR 91	1,103	25.17	142	22.7	32.2	0.16	15.56	20	0.31	4	4	0.02	0.28	1.53
MAY 91	1,137	25.95	186	22.0	31.9	0.59	15.07	21	0.17	4	4	0.03	0.14	1.47
JUN 91	1,171	26.72	167	21.0	31.9	0.08	13.72	17	0.03	2	4	0.02	0.03	1.18
JUL 91	1,210	27.36	160	20.1	31.6	0.06	13.85	13	0.08	3	4	0.03	0.03	1.01
AVERAGE	1,169	26.67	160	21.7	32.1	0.48	14.02	19	0.53	4	5	0.16	0.29	1.78

7.5.1 Characteristics and Properties of Methanol

Methanol (CH₃OH), which is known by a variety of names—such as methyl alcohol, methyl hydrate carbinol, and wood alcohol—is a colorless liquid and is noncorrosive except to aluminum, lead, and some natural rubbers at normal atmospheric temperature. It is normally supplied pure (99.90 percent). Key properties of methanol are listed in Table 7-16. Additional data are available in the literature (34,35). Material safety data sheets (MSDSs) from chemical companies can be extremely useful sources of information about the properties, use,

Table 7-16. Key Properties of Methanol

Density	0.7913 g/mL @ 20°C (6.59 lb/gal)
Vapor density (air = 1.00)	1.105 @ 15°C
Vapor pressure	
@ 0°C	29 mm Hg
@ 10°C	52 mm Hg
@ 20°C	96 mm Hg
@ 30°C	159 mm Hg
@ 40°C	258 mm Hg
@ 50°C	410 mm Hg
Solubility	Miscible in all proportions with water
Viscosity @ 20°C	0.614 cps
Evaporation rate (butyl acetate = I)	4.6
Combustible limits, percent by volume in air at standard temperature and pressure	7.3 to 36
Flash point (tag open cup)	16°C (61°F)
Flash point (tag closed cup)	12°C (54°F)

handling requirements, and hazards associated with methanol; this should be made available to operations personnel.

Taken internally, methanol is highly toxic. It also is harmful if the vapors are inhaled or if it comes in contact with skin for prolonged or repeated exposure. Fire and explosion, however, are the primary dangers associated with methanol. Again, individuals involved in handling methanol should be made aware of these hazards. Federal, state, and local regulations for safety should be posted, along with supplier information.

7.5.2 Regulations and Standards

The shipping, unloading, storage, and handling of any flammable chemical, including methanol, are governed by a number of stringent requirements. These include federal regulations imposed by the Department of Transportation (DOT) and by the Occupational Safety and Health Act

(OSHA); state safety orders and codes; municipal ordinances; guidelines issued by independent associations such as the National Fire Protection Association (NFPA) and the Manufacturing Chemists Association (MCA); and precautions imposed by insurance companies. These requirements must be studied before methanol facilities are designed, and all must subsequently be followed.

7.5.3 Delivery and Unloading

Methanol is shipped in 208 L (55 gal) metal drums, tank wagons, tank trucks, and tank railcars. Other methods of shipping, not discussed at length here, are barge, metal drums (smaller than 208 L [55 gal]), and glass and metal cans. Tank wagons normally hold 3,785–15,140 L (1,000–4,000 gal), tank trucks 15,140–34,060 L (4,000–9,000 gal), and tank railcars 22,700–37,850 L (6,000–10,000 gal). Tank railcars and tank trucks represent the most economical shipping mode for most facilities. However, for pilot work and small plants, 208 L (55 gal) metal drums may be appropriate. Because methanol is classified by the DOT as a flammable liquid, all shipping containers must be approved and labeled in accordance with applicable DOT regulations. Additionally, methanol fumes must be properly vented during transport.

The recommended method of unloading methanol from any container is pumping. Some barges and tank wagons have their own pumps for unloading. Tank railcars and trucks can be unloaded from the top or bottom, and the methanol can be pumped or conveyed by gravity or syphoning. The preferred method of unloading is pumping from the top via an eductor tube. Because of the increased likelihood of spillage when unloading from the bottom, railcars or trucks must be approved for bottom unloading. They must incorporate valving that is approved by the Association of American Railroads (AAR) and that meets DOT requirements. This valving helps contain the product by safely controlling flow. Additional precautions such as fusible link valves and excess flow valves may be used.

Air pressurization of the tank (air padding) must never be used for methanol unloading. However, top unloading using a displacement method with an inert gas such as carbon dioxide or nitrogen for padding may be permissible if the unloading procedures follow the chemical supplier's specifications exactly. Unloading procedures should be validated against current supplier information and pertinent regulations.

General requirements for the design of unloading facilities for methanol apply to both tank railcars and trucks. The unloading area should be located away from the plant's principal traffic areas. Also, all storage facilities should be located outside because of the fire hazard, and all equipment in the vapor area must be explosion-proof—Class I, Group D, Division 1 or 2—per the National Electrical Code. Tools used in unloading should be of the nonspark-

ing type. Unloading should occur during daylight hours because the safety and lighting requirements for night operation are particularly extensive. Ample fire extinguishers, safety blankets, deluge showers, eye washes, no-smoking signs, and unloading-area signage are also required.

If top unloading is utilized, approach platforms are required for access to the top of the tank. In all unloading setups, all equipment must be grounded, including the shipping vessel, interconnecting piping, pumps, and approach platforms. Also, bonding jumpers must be used to provide a favorable continuous system, and the grounding system must be checked routinely.

Static electricity buildup must be minimized because a spark discharge can easily start a fire or cause an explosion. Refer to National Fire Protection Association codes for ways to prevent excessive static electricity.

Truck unloading areas must faciltate truck maneuverability, upon both entering and leaving the area. When the number of individual unloading spots is determined, factors such as frequency of use, amount of space, and ability to unload several truck simultaneously should be considered. To facilitate truck traffic flow, it is best to have parallel unloading setup areas such that trucks can drive straight through.

Rail unloading requires additional considerations. Unloading areas must have derailers or a closed switch at a minimum of one car length away from the car being unloaded. A primary issue in rail unloading is who uncouples and parks the cars—the plant or the railroad. It is preferable to have private sidings so the railroad can drop off or pick up cars at any time without disrupting plant operations. Also, the cost of having the railroad uncouple and park cars at night, over weekends, or on holidays can be high. In addition, because the railroad usually cannot guarantee arrival time of shipments, safety provisions and lighting must be provided for night operations. Two sidings should be provided—one for empty cars and one for full cars. It is recommended that sufficient storage be provided to accommodate the contents of a full rail car. The cars can be maneuvered by plant personnel using railcar movers that operate both on rails or on a roadway surface. In the case of short distances, winches can be used. By paving the railroad yard area, both truck and rail unloading can be practiced. This is advantageous in the event of strikes affecting either kind of transport.

Although unloading equipment is typically made of steel, many other materials (except aluminum) are acceptable, providing they can withstand pressure loads and are completely grounded. Pumps must be of the nonsparking type, such as bronze-fitted steel pumps with bronze impellers. Many materials are compatible with methanol, so seals and gaskets can be constructed from common materials. Pumps may be either centrifugal or the positive-

displacement types; however, positive displacement pumps must have relief valves. Because of the widely varying heads encountered during unloading, a pump should be carefully selected. Piping should have as few joints as possible and should have a Schedule 40 rating as a minimum. Splash guards at joints may be desired in traffic areas. Valves may be of the gate-, plug-, or diaphragm-type, and may be made of iron or steel with bronze trim. Neoprene can be used for the plugs in the plug valve or for the diaphragm in the diaphragm valves. Some refineries on the West Coast have adopted a standard of using cast steel valves in all lines carrying flammable materials, to prevent damage during a fire. Couplings must be leak-proof. It is preferable to have a valve next to the coupling to limit material leakage and waste during disconnection. If flexible hose connections are used, a coupling with an integral valve can be used. A strainer should be used ahead of any pumping or storage equipment.

Care must be exercised to avoid overfilling the storage vessel. A high-level alarm and pump shutoff should be used. Because of the cost of methanol, it may be desirable to have a flow meter in the unloading piping to monitor flow level. All vessels must be vented during unloading or loading.

7.5.4 Storage

In order to accommodate delays in methanol delivery, facilities should have the capacity to store a two- to four-week supply. The volume of storage will be determined by various site and cost requirements; however, storage of less than a two-week supply would provide too small a buffer for expected delivery delays and strikes. Tank truck deliveries require in-plant storage; with rail deliveries, the railcars can be used for storage, though charges (demurrage) are levied by the carriers for time on site in excess of a fixed time. For small plants, demurrage may be cost effective; however, carriers may have a time limit on the loaning of cars or have excessive demurrage charges.

Although methanol may be stored in vertical or horizontal aboveground tanks, which can be located inside or outside, it is strongly recommended that all methanol equipment and tanks be located outside to minimize dangers associated with the possibility of explosions caused by methanol fumes. If interior storage is required, refer to NFPA codes for detailed requirements. An exception to this rule involves drums that, if not stored indoors, must be shaded from direct sunlight or constantly sprinkled with water.

Methanol tanks should be laid out in accordance with the requirements of the NFPA. There should also be a dike around each aboveground tank or group of tanks to contain 125 percent of the largest tank's volume in case of a rupture or fire. If the tanks are not made of steel, care

245

must be taken to ensure that a fire will not rupture the entire group of tanks, thereby causing methanol to overflow the dike. Fire protection is critical, especially when the tanks are near other structures. For large volumes of methanol storage, low-expansion alcohol-type foam is used for fire extinguishing. For small fires, dry chemical or carbon dioxide extinguishers can be used. Temperature detectors may be used for sensing fire and initiating automatic release of foam. Water should not be used.

Storage tanks are normally made of steel, but most comparable materials are satisfactory (except for aluminum and lead). Tank size depends on the required capacities and any size limitations imposed by the tank material. Piping, valves, and other components should be as described in Section 7.5.3. Tank fittings should include:

- An inlet with dip tube to prevent splash and static electricity.

- An anti-siphon valve or hole on the inlet to prevent back siphonage.

- A vent pipe with a pressure-vacuum relief valve (36) with flame arrester.

- An outlet connection.

- A drain connection.

- Various openings for depth gauges, sample points, and level switches.

Tanks should also have manholes for access. Extreme corrosion will take place if the tank is drained dry and not cleaned. The tank must also be grounded. Because of increasing air pollution requirements, venting must be controlled by conservation vents or by maintaining a slight negative pressure in the tank using a small ejector.

To maintain the proper quantity of materials, a diaphragm level sensor or float should be used. Low- and high-level alarms are needed to protect against overfilling and against pumping out settled material from the bottom of the tank. The high-level alarm should be separate from the tank sensors for a fail-safe design.

7.5.5 Transfer and Feed

Methanol must be controlled during transfer from the storage vessel to the point of feed, which is accomplished by gravity-feed or with pumps. Generally, pumps are easier to control. Transfer pumps should always have positive suction pressure and should be protected by a strainer. As with all methanol use, it is desirable to mount all equipment outside. There are two basic pumping arrangements that can be used: 1) positive displacement chemical feed pumps with adjustable stroke, speed, or both, where flow rate can be measured continuously or calibrated periodically, and the flow rate is based on speed and stroke settings; or 2) centrifugal or regenerative turbine pumps with variable-speed drives controlled by a flow meter. Each arrangement has its own particular limitations and must be considered on an individual basis for each installation. However, positive displacement pumps are more common in chemical feed applications because they are more accurate and less dependent on discharge pressure.

All pumps and piping should be uniform, as noted in Section 7.5.3. All piping should be tested for 1.5 times the maximum system pressure for 30 min, with zero leakage.

7.5.6 Feed Control

Because methanol is expensive and an overdose can result in a high effluent BOD, it is essential to pace the methanol feed accurately with the oxidized nitrogen load. Simply pacing methanol dose against plant flow will lead to inaccuracies since this approach does not account for daily and diurnal variations in the nitrate concentration.
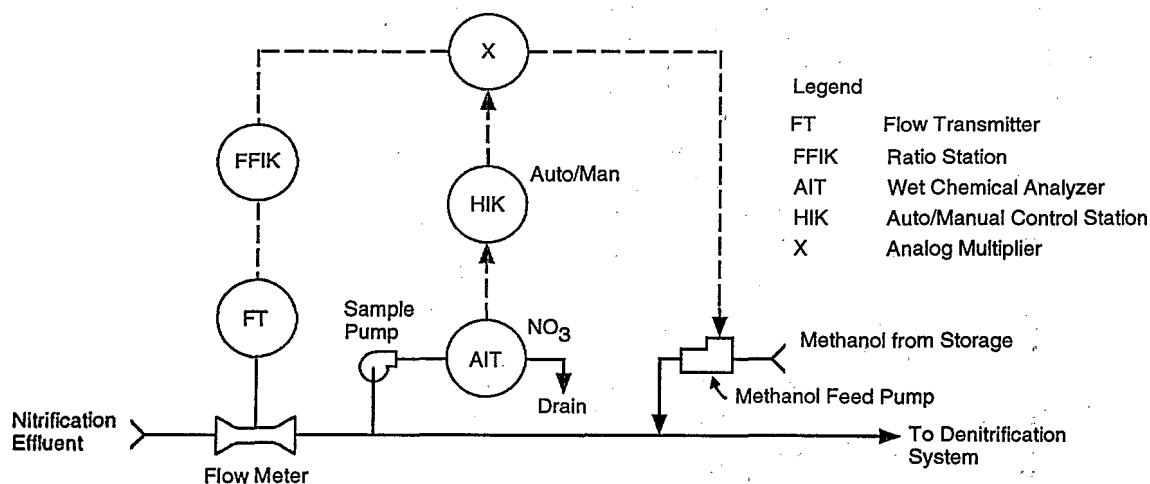

Figure 7-18. Schematic of automatic feed forward control system for methanol pacing.

There are two methods that can be used to control methanol dose: automatic feed forward control and manual control. Manual control is more commonly used.

Feed forward control using plant flow and nitrification system effluent nitrate is shown in Figure 7-18. Feed ratio is approximately three parts methanol per one part nitrate nitrogen by weight (see Chapter 4). This control method, which rarely proves successful and is not commonly used, requires continuous online measurement of nitrate using an online, automated, wet chemistry analyzer or frequent grab samples for methanol measurement.

In the online measurement system, the AIT output is proportional to nitrate concentration in the nitrification effluent. The manual control station (HIK) provides means to select either the analyzer output or to enter a manual concentration value in case of analyzer failure. The output of HIK is multiplied by a signal proportional to flow from the ratio stations (FFIK) to obtain a signal proportional to the required methanol flow ratio. This signal may then be fed to a chemical proportioning pump, as shown in the figure, or may be the setpoint of a flow control loop. FFIK provides a means to adjust the methanol feed ratio. The dependability of this control procedure is predicated on the reliability of the automated wet chemical analyzer. These analyzers require very careful routine maintenance and calibration and are not commonly used.

The most common technique for feed control is the manual method in which operators dose methanol until a hydrogen sulfide smell is recognized (13). However, this method can make it difficult to control the chlorination system because of intermediate formation of nitrite-nitrogen.

7.5.7 Removal of Excess Methanol

Unless specific measures are taken to provide for methanol removal, addition of methanol above stoichiometric requirements (see Chapter 4) will cause methanol to appear in the denitrification process effluent. In one reported instance, a methanol overdose caused an effluent BOD_5 of 106 mg/L (1). Placing total reliance on the methanol feed control system to prevent methanol overdoses may be unrealistic in small plants, where a trained technician's attention can be expected to be infrequent. The provision of a reliable methanol control system and a methanol removal system as a backup should allow nearly fail-safe operation in terms of preventing effluents containing high levels of organics.

A modification of the suspended growth denitrification process, geared to prevent methanol bleedthrough, is shown in Figure 7-1. After denitrification, mixed liquor passes to an aerated stabilization tank where facultative organisms "switch over" from using nitrate to using DO and oxidize any remaining methanol. While there is some uncertainty about the length of time required for facultative bacteria to make the switch and complete methanol

oxidation, it has been reported that 30 min of aeration in an aerated stabilization tank at a Burlington, Ontario, pilot plant was insufficient, as high effluent methanol values were periodically observed in the system (1). However, other plants, including River Oaks in Florida, have reported 30–60 min to be acceptable (4). A period of about 48 min has been found to be sufficient for methanol oxidation (1). Therefore, a 1-hr aeration is adequate on the basis of experience to date. There must be a sufficient population of microorganisms acclimated to the presence of methanol to oxidize the excess methanol. A sudden shock load of methanol will not be oxidized readily because the microorganisms will not be acclimated to the high dose. Anaerobic systems for methanol oxidation are not recommended as a rule because they require even more time for acclimation. Further details concerning solids-liquid separation are presented in Section 7.2.2.

In attached growth denitrification systems, the provision of an aeration basin after the denitrification column does not ensure oxidation of excess methanol. This is because the mass of facultative organisms in the column's effluent is not sufficient to oxidize the carbon biologically. The mass is insufficient because denitrifying organisms are retained in the column's media; only a few pass into the column's effluent. Therefore, systems that remove excess methanol are ineffective in attached growth systems. Careful monitoring and alert operators are necessary to ensure that excess methanol is not discharged.

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Springfield, VA 22161
703-487-4650

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Chapter 8
Design Considerations for Single-Sludge Nitrification-Denitrification Processes

8.1 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. 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 biological sludge. 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, oxidation ditches, sequencing batch reactors, and cyclical aeration systems. Some manufacturers have obtained patent rights for various types of systems or portions of systems and impose a licensing fee for their use. Nonproprietary versions of some of these systems have been developed by other manufacturers. Manufacturers of proprietary systems generally offer performance guarantees and design assistance.

Single-sludge systems are available with a variety of design layouts, reactor configurations, inlet feed arrangements, compartmentalization or baffling, mixing processes, return sludge 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.

Single-sludge systems offer several advantages over multiple-sludge systems or separate-stage systems. 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. Factors to consider compared to separate sludge/stage systems are space availability, reduction in alkalinity consumption, use of wastewater carbon as a carbon source for denitrification in lieu of methanol and lower oxygen requirements.

Potential limitations or disadvantages to consider compared to separate sludge/stage systems include greater sensitivity to toxicity or inhibition without a separate upstream biological treatment step, lower nitrogen removal efficiency, higher energy usage (compared to separate stage), larger volumes of reactor tankage, and greater site requirements. These potential disadvantages are site specific and all or none may apply to a particular situation.

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. Single-sludge systems can more readily be used in retrofitting existing activated sludge plants for nitrogen removal, particularly if the plant has excess capacity. 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.

This chapter provides an overview of generic types and classification of single-sludge processes and describes

available single-sludge systems, including operational parameters, typical performance and process design features, considerations for selection and design of single-sludge systems, and process design examples.

8.2 Classification of Single-Sludge Processes

8.2.1 Classification System

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 microorganisms (i.e., by depleting the cell's mass) or by an exogenous source such as the BOD of the influent wastewater 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, or 3) conveying the flow sequentially through alternating anoxic/aerobic zones. Since denitrification cannot occur without nitrification occurring first, systems are designed and sized to completely nitrify the oxidizable influent TKN. Thus, conventional parameters such as F/M ratio, retention time, oxygen transfer rate, and solids retention time (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. The classifications are presented in approximate order of familiarity and complexity.

1. *Multistage.* 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 class can be further subdivided into:

   a. Single Anoxic Zone (Section 8.2.2). 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. Configurations utilizing endogenous carbon denitrification are not generally used at full scale. 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.

   b. Multiple Anoxic Zones (Section 8.2.3). 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: 1) recycling nitrified mixed liquor to an antecedent anoxic zone, 2) stepfeeding raw wastewater or primary effluent to an anoxic zone containing nitrates, or 3) 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.

2. *Multiphase/Cyclical Aeration* (Section 8.2.4). 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 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.

3. *Oxidation Ditches* (Section 8.2.5). 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.

4. *Sequencing Batch Reactors (SBRs)* (Section 8.2.6). 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.

Several manufacturers of proprietary processes can provide various modifications to achieve nitrogen or phosphorus removal, or both. In cases where more than one process is available, the process discussion will be limited to the process designed exclusively for nitrogen control. Examples of processes with multiple variations include the Bardenpho (four- or five-stage), Kruger, and Schreiber processes. Although the five-stage Bardenpho process is more common than the four-stage, only the latter will be discussed in detail since it is designed exclusively for nitrogen removal.

8.2.2 Single Anoxic Zone

8.2.2.1 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 (1), but a concurrent system was developed by Ludzack and Ettinger (2). These two systems are presented schematically in Figures 8-1 and 8-2, respectively. The difference between these two systems is related to the carbon source for the denitrifying population. The Wuhrmann 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 (3) was able to demonstrate 88-percent TN 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
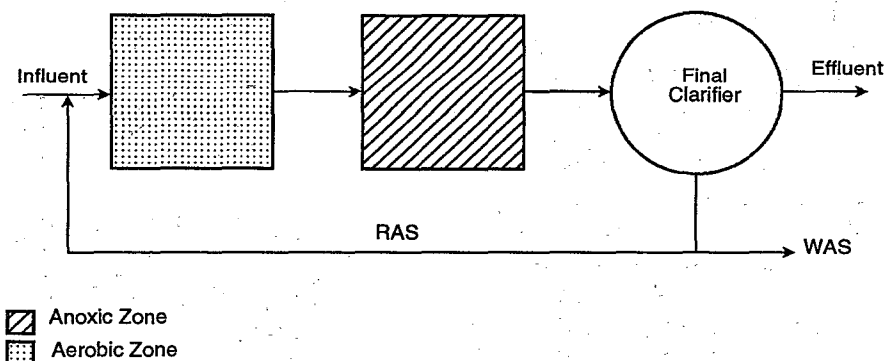

Anoxic Zone
Aerobic Zone

Figure 8-1. Wuhrmann process.
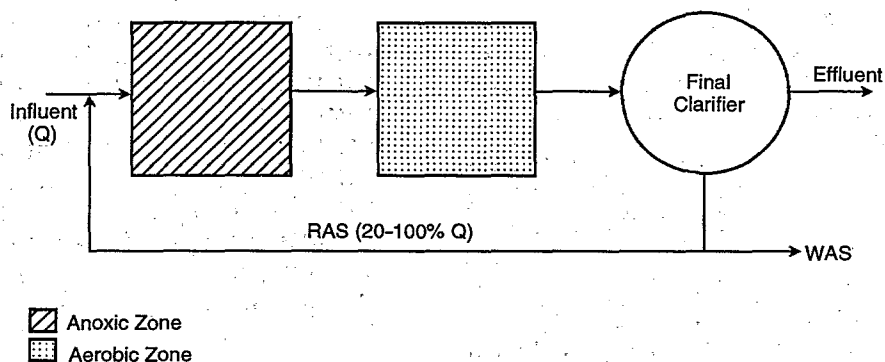

Anoxic Zone
Aerobic Zone

Figure 8-2. Ludzack-Ettinger process.

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 (4) improved the Ludzack-Ettinger process by providing an additional internal MLSS 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 percent were achieved. The MLE, by Barnard, schematically illustrated in Figure 8-3, was not extensively implemented at full scale, but was the progenitor of proprietary configurations, such as the A²/O, Bardenpho, UCT, and VIP. Variations of the MLE process design have been investigated by German and Japanese researchers (5-7). Schreiber and Menzel (6) 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 A²/O (anaerobic/anoxic/oxic) 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 zone. 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 (8). A discussion of selectors is provided in Section 8.4.7. A schematic diagram of the A²/O process is presented in Figure 8-4. The A²/O process is now marketed and licensed by Kruger Company.

The UCT process was developed at the University of Capetown in South Africa to surmount one of the inherent limitations of the MLE and A²/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 Figure 8-5. The purpose of these modifications is to denitrify nitrates returned by the RAS (reutrn 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 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.

8.2.2.2 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 A²/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.

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 6. In summary, the sizing of the aerobic zone should consist of the following steps:

1. Select the design aerobic θ_c^d.

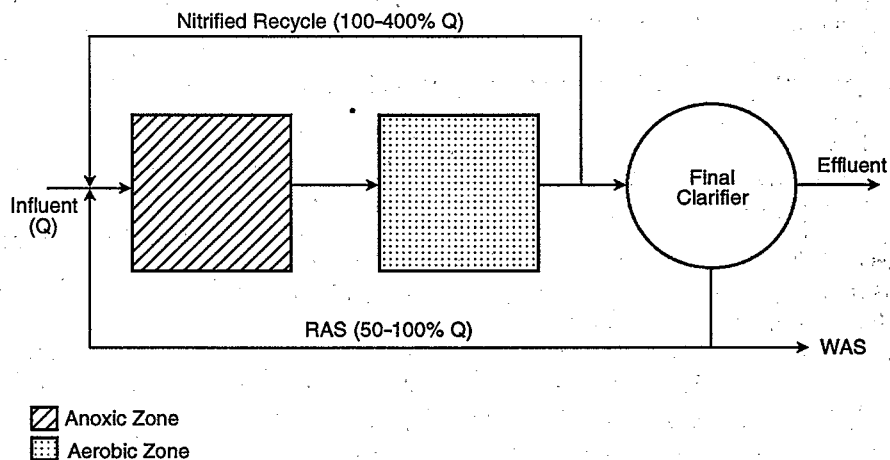2. Calculate secondary sludge production.

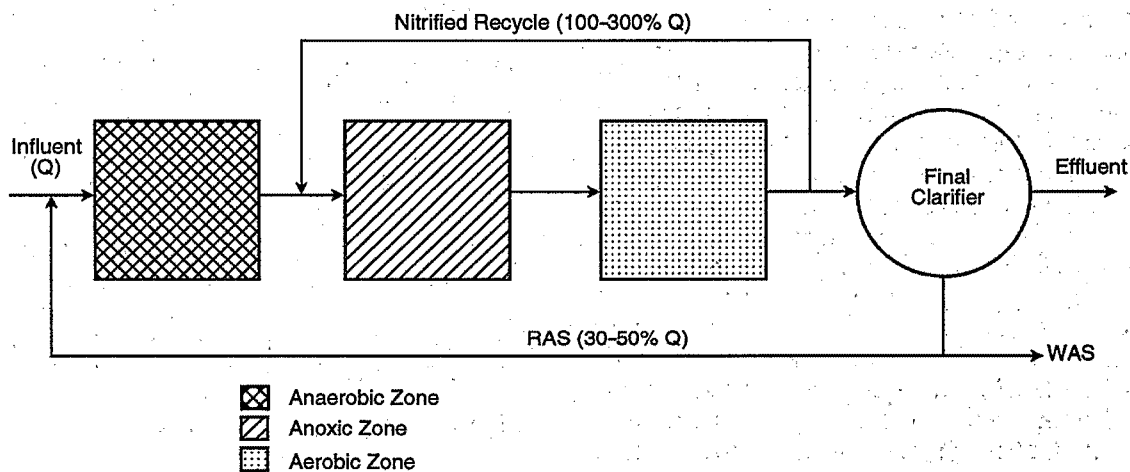Figure 8-3. Modified Ludzack-Ettinger process.

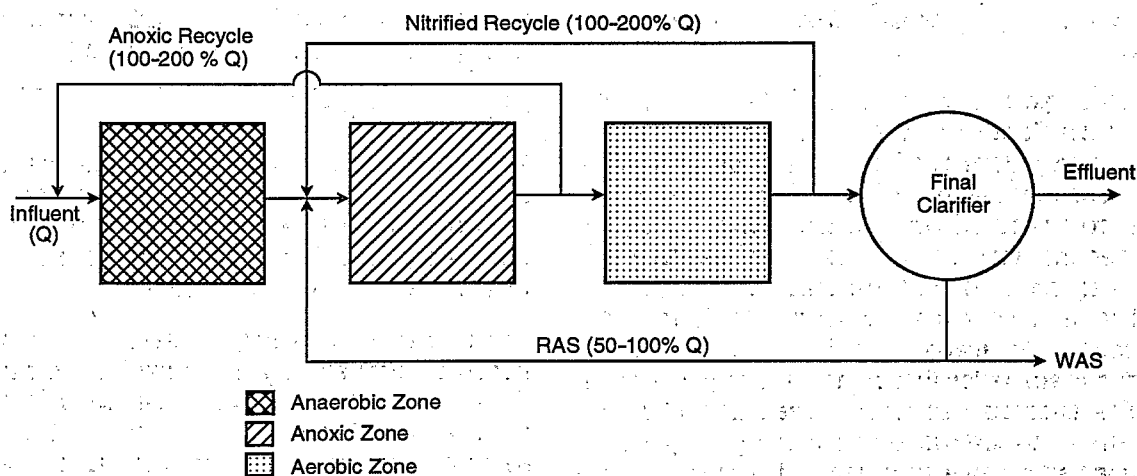Figure 8-4. A^2/O process with nitrification-denitrification.

Figure 8-5. University of Capetown (UCT) process.

3. Calculate the required aerobic zone solids inventory based on θ_c^d.

4. Determine tank volume based on the solids inventory, settling properties, peaking factors, and the design MLSS.

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. From mass balance considerations, the required combination of internal mixed liquor recycle rate and return activated sludge recycle rate can be calculated using Equation 8-1:

$$[(TKN_{ox})/(Q \; C_e)] - 1 = RAS + I \qquad (8-1)$$

where:

TKN_{ox} = total mass of oxidizable TKN (nitrates) produced in the aerobic reactor, g/d

= influent TKN − effluent TKN − TKN assimilated in waste sludge, g/d

Q = plant influent flow rate, m³/d

C_e = desired effluent nitrate concentration, mg/L

I = nitrified internal recycle ratio of recycle rate to plant influent flow rate

RAS = return activated sludge recycle ratio of return sludge rate to plant influent flow rate

Typically, the return sludge flow rate (Q_{RAS}, m³/d) will be calculated during the aerobic zone BOD/nitrification design step as the recycle rate necessary to maintain the design MLSS leaving only the required internal recycle rate (Q_I, m³/d) to be calculated. Equation 8-1 may be algebraically manipulated to express Q_I and Q_{RAS} in terms of a desired percent removal, as provided in Equation 8-2:

$$I + RAS = (\% \; Removal) \div (1 - \% \; Removal) \qquad (8-2)$$

where % removal is expressed as a decimal

The anoxic zone should be sized to denitrify the nitrates recycled to it. The mass of nitrates recycled can be simply expressed as:

$$(Q) \; (C_e) \; (I + RAS) = \text{nitrates recycled to anoxic zone, g/d} \qquad (8-3)$$

The required anoxic solids inventory can be calculated based on the selected SDNR (specific denitrification rate), from which the anoxic volume can be derived. Typical design criteria for single anoxic zone single-sludge nitrification-denitrification processes are provided in Table 8-1.

The design criteria for the A²/O process presented in Table 8-1 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, which represents the only full-scale application of the VIP/UCT process. The VIP criteria differ from the A²/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

Table 8-1. Typical Design Criteria for Single Anoxic Zone Predenitrification Systems

Parameter	A²/O	VIP/UCT	Generic Single Anoxic Zone
MLSS, mg/L[a]	3,000–5,000	1,500–3,000	1,500–4,000
HRT, hr			
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, d	5–10	5–10	5–10
F/M, g BOD₅ applied/g MLVSS/d	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

[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] A²/O and MLE do not incorporate an anoxic recycle. Anoxic recycle is not required for systems that do not remove phosphorus.

removal by providing two internal recycles. This modification affords a greater total recycle of nitrates for denitrification without affecting phosphorus removal processes. The A²/O process is generally operated at a higher MLSS than the VIP and at a lower RAS rate. The lower RAS rates in the A²/O process are required to ensure that the anaerobic selector is not overloaded with nitrates, which would adversely affect phosphorus removal.

8.2.2.3 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.

As demonstrated in Equation 8-2, 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 ≤5 mg/L, or at facilities where >80 percent TN removal is required. Return sludge rates are generally limited to 100 percent 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.

The theoretical efficiency of a single anoxic reactor system may be derived from a mass balance analysis on nitrates for a given reactor train, as illustrated in Figure 8-6. The quantity of total nitrates produced in the aerobic zone, assuming complete nitrification of the available TKN, is the oxidizable TKN or TKN_{ox}. A fraction of these nitrates will be recycled to the anoxic zone via the internal MLSS and RAS recycles and the remainder will be passed to the effluent. The recycled fraction (fNO_3) may be quantified mathematically as:

$$fNO_3 = (Q_I + Q_{RAS}) \div (Q + Q_I + Q_{RAS}) \qquad (8-4)$$

where Q, Q_I, and Q_{RAS} are the plant influent, internal recycle, and return sludge flow rates, respectively, expressed in consistent units of volume per time unit.

For convenience, the recycle flows may be expressed as a multiple of the influent plant flow, yielding:

Figure 8-6. MLE mass balance schematic.

255

$$fNO_3 = (I + RAS) \div (1 + I + RAS) \qquad (8\text{-}5)$$

where:

$$I = Q_i/Q$$
$$RAS = Q_{RAS}/Q$$

The fraction of nitrates removed shown in Equation 8-5 for a single anoxic system is equivalent to the percent removal. Figure 8-7 illustrates the relationship between percent removal and total recycle ratio. The minimum effluent nitrate concentration (C_e) will depend on the nitrate that is not recycled to the anoxic tank, expressed as:

$$C_e = (1 - fNO_3)(TKN_{ox} \div Q) \qquad (8\text{-}6)$$

or, expressed in other terms as:

$$C_e = (TKN_{ox}) \div [(Q)(1 + I + RAS)] \qquad (8\text{-}7)$$

Thus, the mass fraction of the total nitrate formed in the aerobic zone that is removed in a predenitrification system is a function of the internal recycle rate from the aerobic zone and of the return sludge rate, as presented in Equation 8-7.

A plot of theoretical oxidizable nitrogen removal rate versus internal recycle for typical return sludge rates (i.e., 50–100 percent) is illustrated in Figure 8-8. The figure demonstrates that the maximum removal efficiency for a single anoxic reactor is 85 percent of oxidizable TN, at realistic recycle rates (i.e., ≤400 percent). These relationships do not consider denitrification that may occur in the final clarifier and within the floc particle in the aerobic zone, as hypothesized (10,1). Consequently, observed nitrate removals may be greater than results predicted by theoretical considerations.

The A²/O process has been implemented at the Largo Wastewater Treatment Plant in Florida. Performance data from that plant are presented in Table 8-2.

The MLE process was also used at Maitland, Ontario, to treat a high-strength industrial wastewater (nitrate = 175 mg/L, NH₄⁺-N = 190 mg/L, BOD₅ = 1,230 mg/L). During optimum conditions, TN removals of 93 percent were obtained (10).

There are no known facilities configured on the UCT process in the United States. The VIP process was designed for 70-percent removal of both TN and TP; however, lower TN removal is anticipated in winter because of lower nitrification rates (9). Theoretically, the UCT and VIP processes should achieve higher degrees of nitrogen removal than the A²/O process as a result of both the extensive use of baffles and compartments, and the denitrification conditioning of the RAS in the anoxic stage. By denitrifying the RAS in the anoxic zone, the nitrate interference on phosphorus removal is alleviated. This affords a higher RAS recycle rate and enhanced nitrogen removal efficiency. Actual performance data for the VIP process in-

dicated that effluent TN was 6.8-11.3 mg/L, and the overall average was <8 mg/L; results comparable to those obtained by the similar A²/O process. Experimental results from the VIP plant are presented in Table 8-2 (11).

Table 8-2. Performance Summary of Single Anoxic Zone Processes

Parameter	A²/O (9) Largo, FL	VIP (11) Pilot Norfolk, VA	MLE Landis, NJ
Q, m³/d	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
NH₄⁺-N inf., mg/L	—	—	—
NH₄⁺-N eff., mg/L	—	1.0	—
NO₃-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

The MLE and A²/O processes can be optimized by varying the internal recycle and return sludge rates. However, if phosphorus removal is required, the return sludge rate must be minimized to control the nitrate level to ensure an anaerobic condition. This nitrate interference of the RAS recycle was cited as a limitation to phosphorus removal and was the impetus for development of the UCT process. If adequate volume exists, the MLE or A²/O can be implemented at an existing facility requiring minor modifications. These modifications may include installation of baffles, mixers, and internal recycle pump, as long as the existing reactor volumes can provide adequate retention times and sufficient aeration capacity is available.

8.2.2.4 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 (12) and concentric circular reactors (5). These optional flow regimes did not appear to offer significant process improvement over the CSTR (6). 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 A²/O) can be sufficiently achieved by a nonrigid baffle system; however,
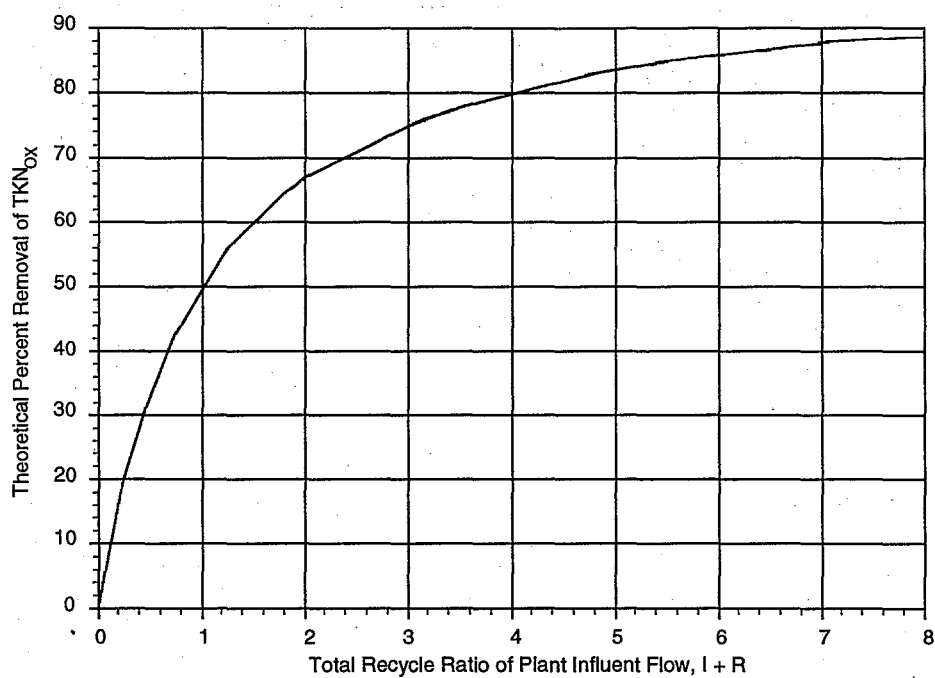
Figure 8-7. Theoretical oxidized TKN removal performance for a single anoxic zone process as a function of total recycle rate.
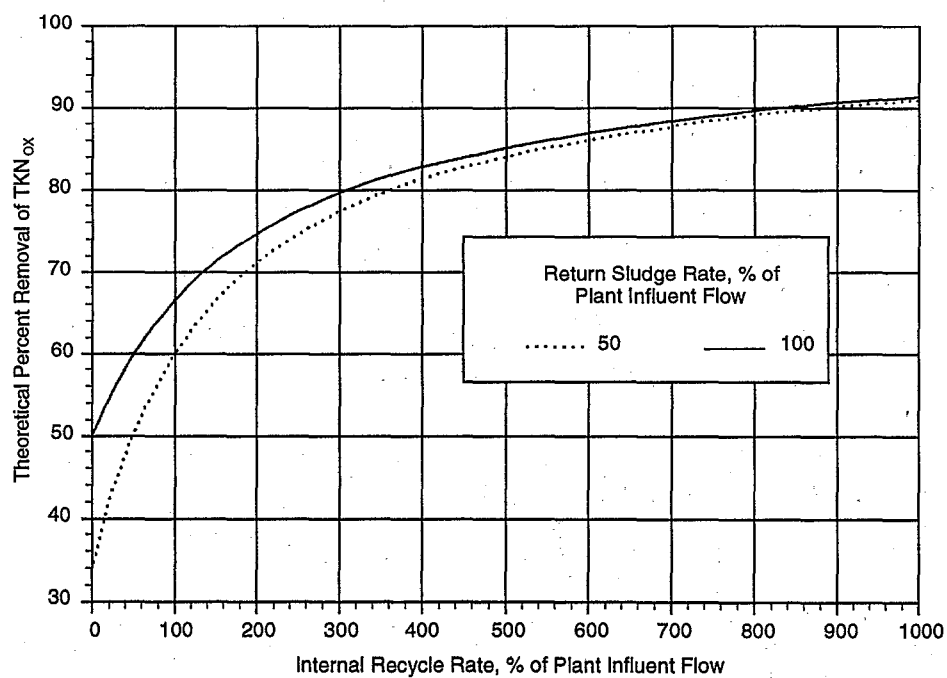

Figure 8-8. Theoretical oxidized TKN removal performance for a single anoxic zone process as a function of internal recycle rate.

257

extensive baffling within each zone as employed in the VIP process did not appear to improve nitrogen removal significantly. An exogenous carbon source for a single anoxic zone system with a predenitrification configuration is normally preferred, since postdenitrification arrangements typically require methanol addition. Monitoring of the denitrification rate should be performed to assess the heterotrophic denitrification capacity for the specific influent COD or BOD. Bypassing the primary effluent to the anoxic reactor was suggested as a means to ensure a higher COD:TKN ratio for improved denitrification performance (6).

Since theoretical nitrogen removal performance considerations are derived based on complete nitrification, TN removal performance is limited by the nitrification efficiency in the aerobic reactor. The chemoautotrophic nitrifiers are susceptible to upsets from shock loadings of BOD and ammonia, and the nitrification rate is highly sensitive to changes in pH, alkalinity, and temperature. The sensitivity of the nitrifiers provides further support for a predenitrification configuration, which can serve as a buffer zone against shock loads to the aerobic reactor. Thus, process optimization will require monitoring of these parameters in the aerobic basin, and adjustments such as alkalinity control and increasing the solids retention time may be necessary.

Design features and considerations for the aerobic zone follow the same procedure as the aeration basin of a nitrification system. A detailed discussion of design for nitrification systems is provided in Chapter 6. Additional design considerations for nitrification-denitrification processes are discussed in Section 8.4.

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 (13). 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, sewage pumps, or vertical turbine pumps directly in the aerobic basin, rather than by constructing a separate dry pit pump gallery.

Both the MLE and A²/O process 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 8-3.

Table 8-3. Monitoring Requirements and Rationale for Single Anoxic Zone Reactors

| Reactor | Ludzack Ettinger, MLE | | A²/O, VIP, UCT | |
	Parameter	Rationale	Parameter	Rationale
Anaerobic	N/A	N/A	DO, Nitrates	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
			NO₃	Inadequate load can cause excess phosphate release
	Qᵢ	Controls NO₃ load	Qᵢ	Controls NO₃ 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

8.2.3 Dual Anoxic Zones

8.2.3.1 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 without methanol addition by placing an endogenous anoxic zone in series after the aerobic zone. While the A²/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 Figure 8-9. 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 Figure 8-9. 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 in Section 8.2.2, 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 sludge (RAS shown in Figure 8-10) and the nitrate recycle separately and also to reduce the NO_3 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 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 multi-anoxic zone process (7) is illustrated in Figure 8-11. 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

4-Stage (Nitrogen Removal) Process

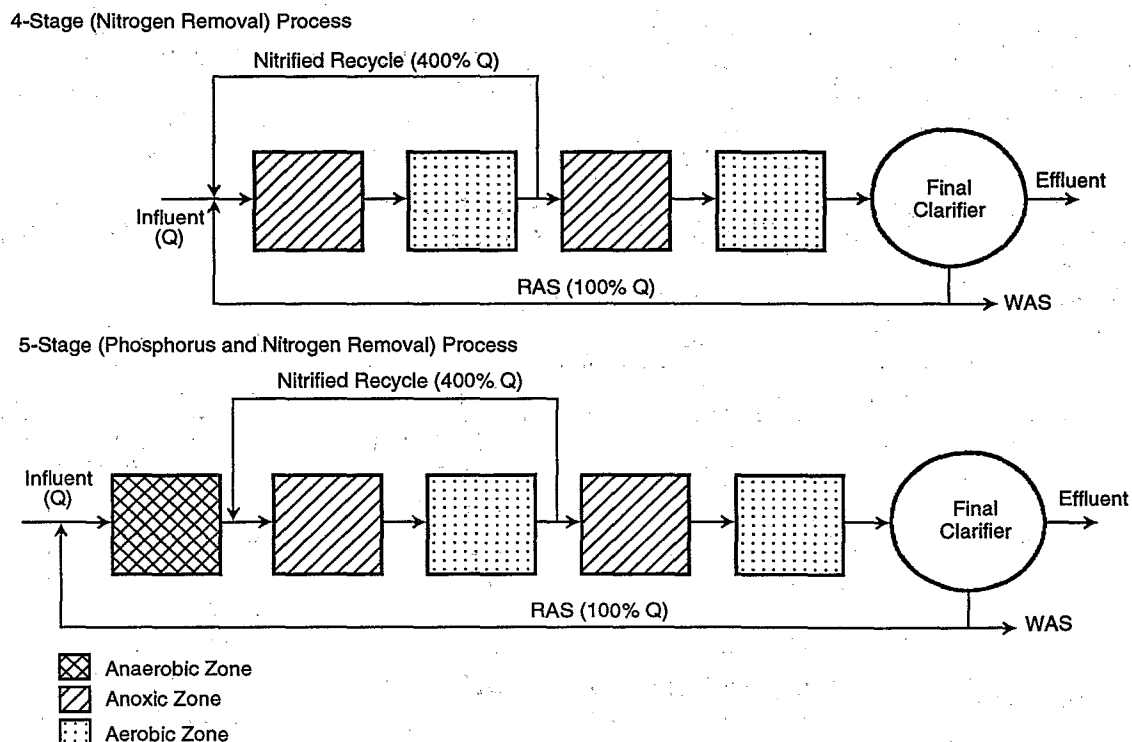5-Stage (Phosphorus and Nitrogen Removal) Process

Figure 8-9.   Bardenpho process.

return. EIMCO administers a one time royalty fee for the process, which can include startup, training, and guarantee of performance (9).

8.2.3.2 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 Section 8.2.2.2 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 sludge 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. Since endogenous denitrification rates are much slower than ex-
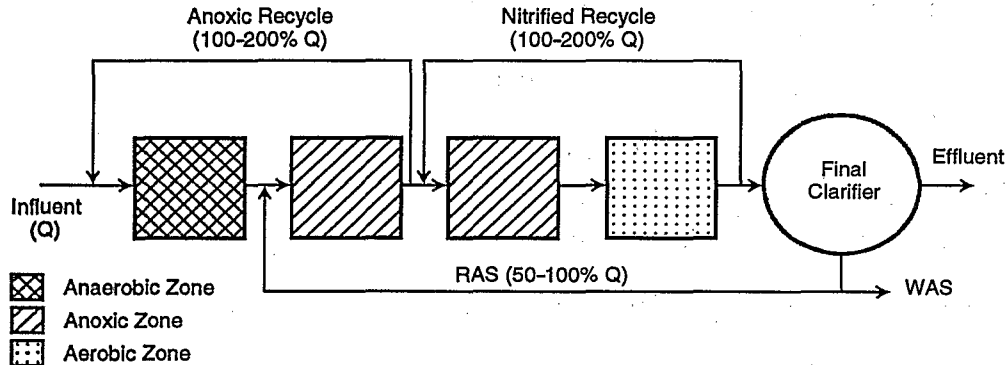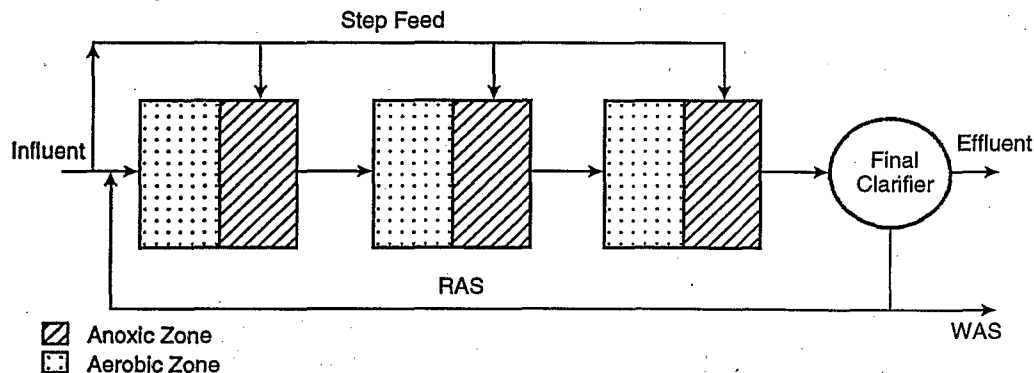

Figure 8-10. Modified UCT process.

Figure 8-11. Multi-anoxic zone with step feed.

ogenous 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 system are presented in Table 8-4.

The modified UCT has two fundamental process differences compared to a five-stage Bardenpho:

1. The Modified UCT is designed to optimize phosphorus removal.

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. 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.

Table 8-4. Typical Design Criteria for Dual Anoxic Zone Systems

Parameter	4-Stage Bardenpho	Modified UCT
F/M, g BOD_5/g MLVSS/d	0.1–0.2	0.1–0.2
θ_c, d	10–40 ·	10–30
MLSS, mg/L	2,000–5,000	2,000–4,000
HRT, hr		
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

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 Figure 8-11, 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.

8.2.3.3 Process Performance

The Bardenpho design has achieved TN effluent concentrations of 3 mg/L and 90-percent removal afforded by the endogenous postdenitrification stage. The Bardenpho process has been used at several plants in the United States. A list of typical performance data is included in Table 8-5.

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.

A mass balance analysis provides insight to the higher nitrogen treatment efficiency obtainable with a Bardenpho process compared to the Modified UCT system. The theoretical nitrate removal efficiency for each system can be described by considering their respective recycle patterns.

The fraction of total nitrates produced in the nitrification tanks that are removed is a function of the internal recycle rate from the aerobic to anoxic zone. This rate is typically 400 percent and 200 percent of the influent flow for the Bardenpho and Modified UCT, respectively. The nitrate fraction recycled from the first aerobic zone to the first anoxic reactor of the Bardenpho is (assuming a typical RAS rate of 100 percent):

Nitrate to first anoxic zone
$$= [(4Q) \div (4Q + 2Q)] \times TKN_{ox} = (2/3) \, TKN_{ox} \quad (8\text{-}8)$$

Thus, one-third of the TKN oxidized in the aerobic zone passes through to the second anoxic zone. If no denitrification is assumed to occur in the second anoxic zone (an unrealistic assumption) then the total amount of nitrates not recycled to the first anoxic zone, one-third TKN_{ox} would be passed to the clarifier. The nitrates then recycled to the anaerobic reactor of a five-stage Bardenpho process at an RAS of 100 percent of the influent, Q, would be:

Nitrates recycled to anaerobic reactor in RAS
$$= (1/3) \, (TKN_{ox}) \, (Q)/(2Q) = (1/6) \, TKN_{ox} \quad (8\text{-}9)$$

The total fraction of nitrates recycled (calculated in Equations 8-8 and 8-9) would be:

Table 8-5. Summary of Bardenpho Plant Operating Data

Plant	Flow, m³/d (mgd)	Influent BOD₅, 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	4,656 (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	2,574 (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

[a] COD
[b] NH₄-N only

Total fraction nitrates recycled

$$= [(2/3) + (1/6)]\ TKN_{ox} = (5/6)\ TKN_{ox} \quad (8\text{-}10)$$

and the fraction of nitrates discharged to the effluent would be:

Nitrates in effluent

$$= [(1) - (5/6)]\ TKN_{ox} = (1/6)\ TKN_{ox} \quad (8\text{-}11)$$

The preceding discussion can be written in general for any combination of internal recycle or return activated sludge recycle rate, or overall percent removal in the second anoxic zone. From these theoretical considerations, a model can be developed to explain the enhanced performance of the dual anoxic zones of the Bardenpho process compared to the Modified UCT or to a single anoxic zone process. The fraction of nitrates recycled in the internal recycle flow to the first anoxic zone (fNO_3) in a Bardenpho is:

$$fNO_3 = (Q_i) \div (Q + Q_i + Q_{RAS}) \quad (8\text{-}12)$$

or, if the recycle rates are expressed in terms of Q, as shown in Figure 8-12.

$$fNO_3 = (I) \div (1 + RAS + I) \quad (8\text{-}13)$$

where:

$I = Q_i/Q$
$RAS = Q_{RAS}/Q$

The fraction of nitrates not recycled but passed to the second anoxic zone is then:

Fraction of nitrates passed to 2nd anoxic zone
$$= (1 - fNO_3) \quad (8\text{-}14)$$

The amount of nitrates removed in the endogenous denitrification reactor can be expressed as a percent removal of the total oxidized nitrogen load passed to the

second anoxic zone. For discussion purposes, the removal from endogenous nitrate respiration expressed as a decimal will be denoted as ER. The ER nitrate removed is itself a function of the nitrate load that can be evaluated from the mechanistic considerations introduced in Chapter 4. The mass of nitrates removed in the second anoxic reactor using these parameters may be expressed as:

Mass of nitrates removed in endogenous anoxic reactor

$$= [ER\ (1 - fNO_3)]\ (TKN_{ox}) \quad (8\text{-}15)$$

The fraction of nitrates passed to the aerobic reactor and final clarifier is simply the difference from unity of the nitrates passed through the endogenous denitrification reactor, expressed as:

Fraction of nitrates passed to clarifier

$$= (1 - ER)\ (1 - fNO_3) \quad (8\text{-}16)$$

The expression for nitrates passed to the effluent is then the portion of remaining nitrates that are not recycled to the anoxic reactor in the RAS. From expressions similar to before, the nitrates recycled in the RAS may be expressed as a recycled fraction of the total flow to the clarifier, or:
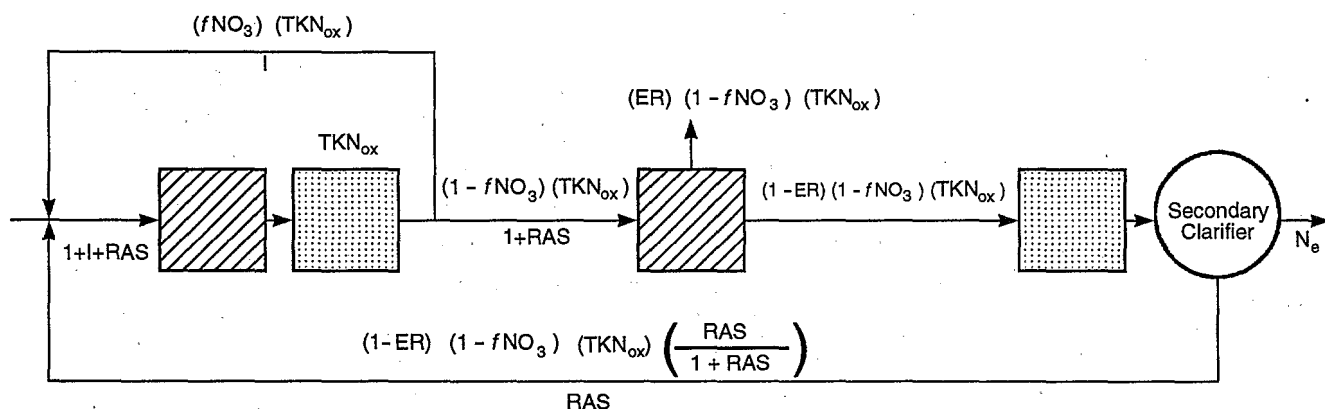
Fraction of nitrates recycled in RAS

$$= [(1 - ER)\ (1 - fNO_3)]\ [(RAS) / (1 + RAS)] \quad (8\text{-}17)$$

The fraction of nitrates passed to the effluent, fN_e, may be calculated by a mass balance relationship as follows:

Nitrates passed to clarifier = nitrates recycled in RAS + nitrates in final effluent

$$(1 - ER)\ (1 - fNO_3)$$
$$= [(1 - ER)\ (1 - fNO_3)]\ [(RAS)/(1 + RAS)] + fN_e \quad (8\text{-}18)$$

(fNO_3) (TKN_{ox})

(ER) $(1 - fNO_3)$ (TKN_{ox})

TKN_{ox}

$(1 - fNO_3)$ (TKN_{ox})

$(1 - ER)$ $(1 - fNO_3)$ (TKN_{ox})

Secondary Clarifier

N_e

$1+I+RAS$

$1+RAS$

$(1-ER)$ $(1-fNO_3)$ (TKN_{ox}) $\left(\dfrac{RAS}{1+RAS}\right)$

RAS

$$\frac{I}{1+RAS+I} = fNO_3 = \frac{Q_I}{Q + Q_{RAS} + Q_I}$$

Nitrates passed to clarifier = Nitrates recycled + N_e

RAS and I expressed as fraction of Q, i.e., I = 1 for 100% recycle

$$(1 - ER)\ (1 - fNO_3)\ (TKN_{ox}) = (1 - ER)\ (1 - fNO_3)\ (TKN_{ox})\ \left(\frac{RAS}{1+RAS}\right) + N_e$$

$$N_e = (1 - ER)\ (1 - fNO_3)\ (TKN_{ox})\ \left(1 - \frac{RAS}{1+RAS}\right)$$

$$\frac{Q_I}{Q} = I \qquad \frac{Q_{RAS}}{Q} = RAS$$

Fraction removed $= 1 - \dfrac{N_e}{TKN_{ox}} = 1 - [(1 - R)(1 - fNO_3)(1 - \frac{RAS}{1+RAS})]$

⧄ Anoxic Zone
▦ Aerobic Zone

Figure 8-12. Bardenpho mass balance schematic.

Solving for fN_e:

$$fN_e = (1 - ER)\ (1 - fNO_3)\ \{1 - [RAS / (1 + RAS)]\}$$

(8-19)

Percent removals as a function of the internal recycle rate (manifested in the fNO_3 term) are displayed in Figure 8-13 for various endogenous removal rates and two RAS rates.

Figure 8-13 demonstrates that under normal conditions, the Bardenpho process can remove 83 percent of the oxidized TKN if no endogenous denitrification is considered. If 50 percent of the nitrates to the second anoxic zone are removed through endogenous nitrate respiration (a conservative estimate), the nitrate removal performance increases to approximately 93 percent.

A similar analysis can be made for the Modified UCT system, illustrated in Figure 8-14. The nitrified internal recycle (I) and RAS recycle directly affect the nitrate removal performance. Assuming all nitrates in the RAS stream are removed in the first anoxic zone, the anoxic recycle (I_A) does not affect denitrification. The expression for the fraction of nitrates recycled in the second anoxic zone of the Modified UCT is the same as the expression for the fraction removed in the first anoxic zone of the Bardenpho process:

$$fNO_3 = I \div (1 + RAS + I) \qquad (8\text{-}20)$$

However, the amount of nitrates passed to the effluent from the final clarifier in the Modified UCT is only affected by the RAS recycle, since there is no additional denitrification afforded by a second anoxic zone sequent to the aerobic zone. This is evident mathematically if ER in the Bardenpho process is set equal to zero, in which case the Bardenpho would achieve the same level of performance as a single anoxic zone system.

Since denitrification in the Modified UCT occurs only in the exogenous denitrification zones, the concentration of nitrates in the discharge from the aerobic reactor will remain constant. Thus, the mass of oxidized TKN in the effluent (or that portion not recycled) can be calculated by simply expressing the combined recycle (I + RAS) as a fraction of the total nitrified flow (1 + I + RAS), as provided in Equation 8-21:

$$N_e = \{1 - [(I + RAS) / (1 + I + RAS)]\}\ TKN_{ox} \qquad (8\text{-}21)$$

The above expression may be algebraically manipulated to verify its equivalence with the expression for N_e derived in Figure 8-14.

Figure 8-15 demonstrates that the maximum feasible percent nitrogen removal of the Modified UCT process is <90 percent.
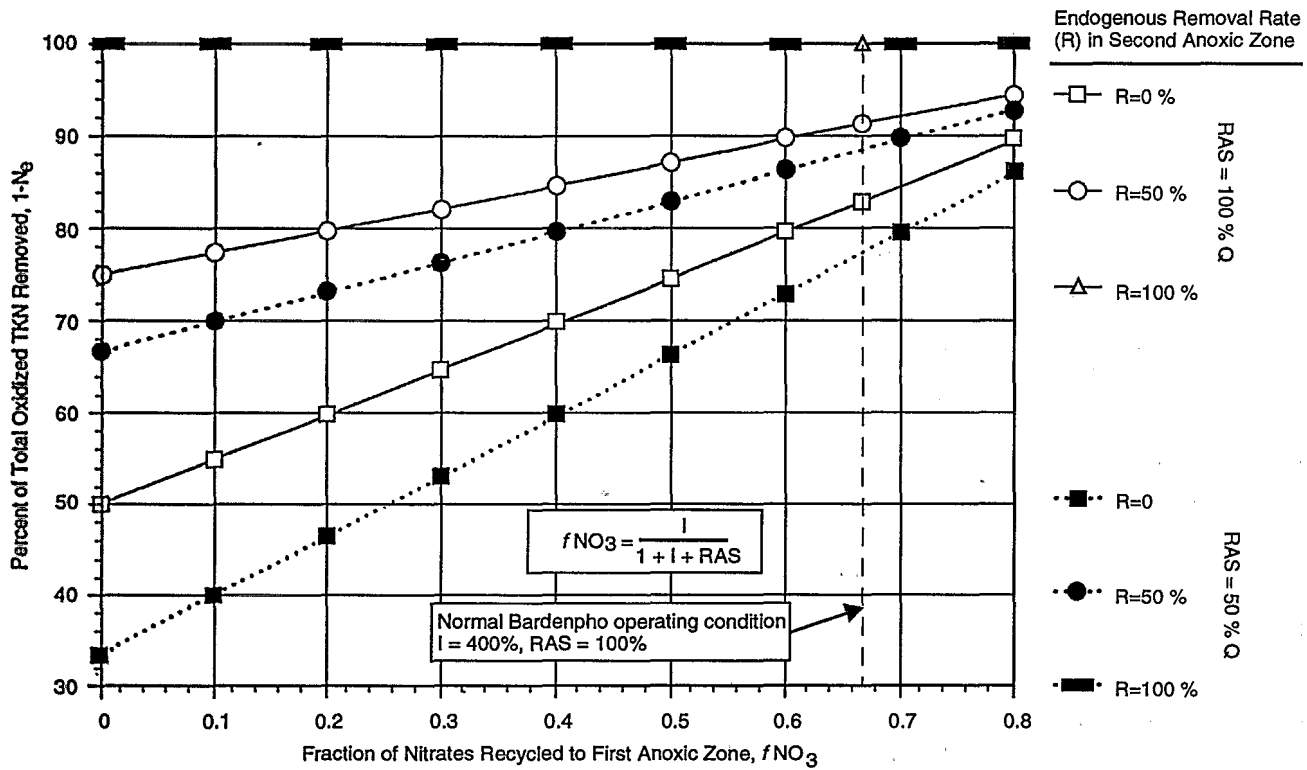
Figure 8-13. Bardenpho process nitrate removal as a function of recycle rates and denitrification performance.
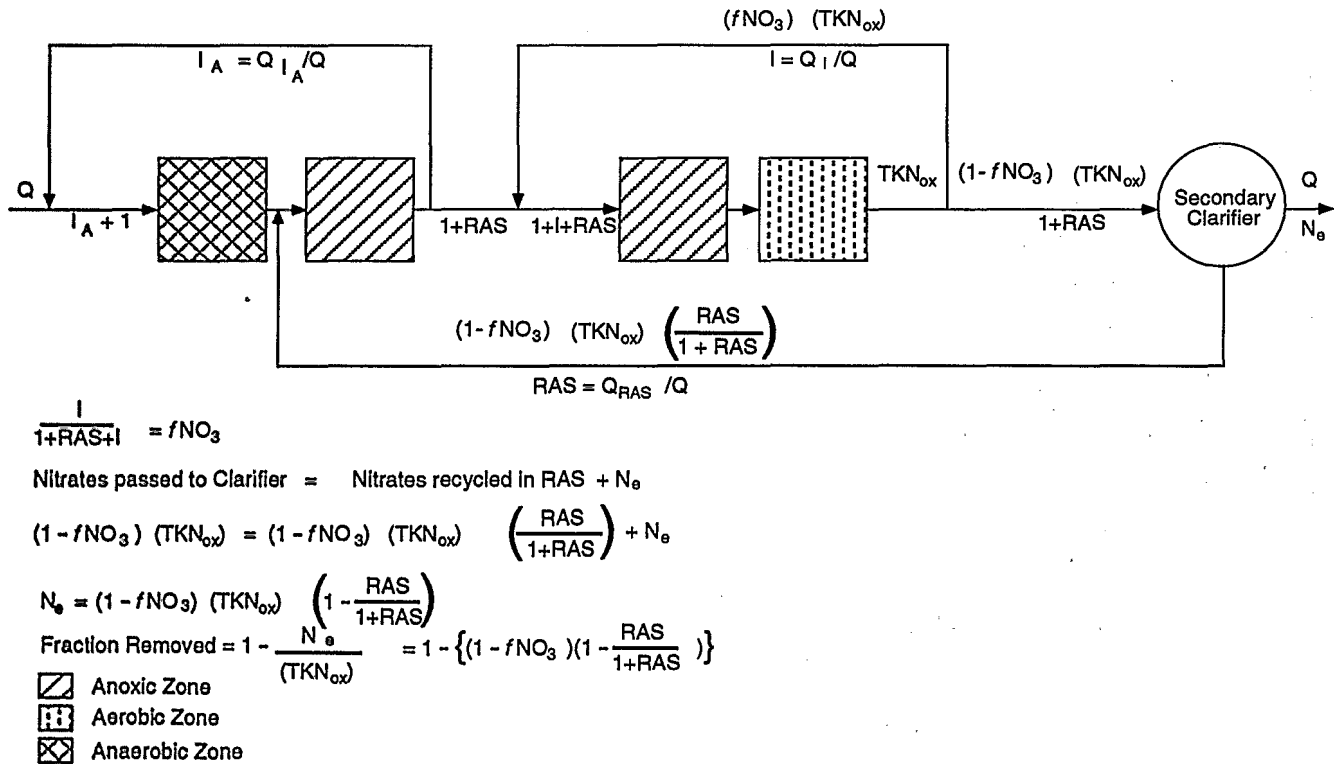

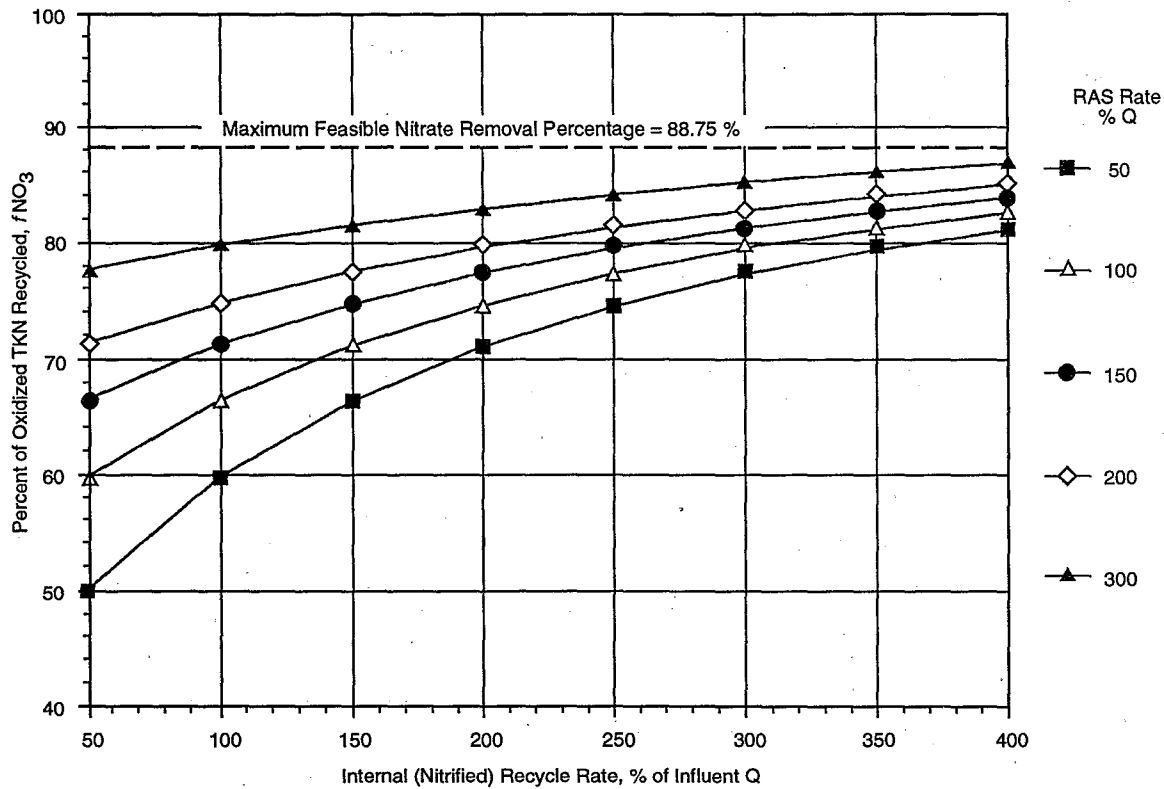**Figure 8-14.** Modified UCT mass balance schematic.

Figure 8-15. Modified UCT process nitrate removal as a function of internal and return sludge recycle rates.

The theoretical mass balance necessary to optimize a staged sequence of aerobic-anoxic zones with step feed can be derived using similar concepts to those used for recycle systems. A definition schematic is presented in Figure 8-16. The influent wastewater mass load of TKN and biodegradable COD is denoted with the subscript zero. The mass of TKN and biodegradable COD to the first aerobic basin is proportional to the step-feed ratio S, such that:

$$TKN_1 = S_A \, TKN_0 \qquad (8\text{-}22)$$

$$COD_1 = S_A \, COD_0 \qquad (8\text{-}23)$$

where S_A is the ratio of the flow diverted to the first anoxic basin (volume/time) to the total process influent flow (Q). In Figure 8-16, $S_A + S_B + S_C + S_D = 1.0$.

If the aeration system is sized to provide for oxidation of the incoming oxidizable fraction of TKN_1, the nitrate load to the first anoxic reactor, $NO_{3(1)}$, can be expressed simply as:

$$NO_{3(1)} = TKN_1 - 0.015 \, COD_1 - N_{ref1} \qquad (8\text{-}24)$$

where $NO_{3(1)}$ is the nitrate mass. The term $0.015 \, COD_1$ accounts for the nitrogen incorporated into cell mass, and N_{ref1} denotes the refractory nitrogen mass that was included with TKN_1.

The COD necessary for denitrification in the first anoxic zone must be provided by step feeding. The mass of

biodegradable COD step-fed into the first anoxic zone is given by:

$$COD_2 = S_B \, COD_0 \qquad (8\text{-}25)$$

A fraction of TKN_0 will also be present in the step feed. This fraction is given by:

$$TKN_2 = S_B \, TKN_0 \qquad (8\text{-}26)$$

Both TKN_2 (including its refractory organic nitrogen component, N_{ref2}) and N_{ref1} will be passed to the second aerobic zone.

The nitrates that are denitrified in the first anoxic zone will consume 2.9 g of COD per g of nitrate denitrified (Chapter 2). Thus, the influent wastewater biodegradable COD passed to the second aerobic zone is:

$$COD_3 = COD_2 - 2.9 \, NO_{3(1)} \qquad (8\text{-}27)$$

The above expressions can be expanded to describe a series of subsequent reactors as a function of the influent wastewater TKN, biodegradable COD and step-feed ratios.

$$\begin{aligned} NO_{3(2)} &= TKN_3 - 0.015 \, COD_3 - N_{ref3} \\ &= TKN_2 - 0.015 \, COD_3 - N_{ref2} \qquad (8\text{-}28) \end{aligned}$$

$$\begin{aligned} NO_{3(3)} &= TKN_5 - 0.015 \, COD_5 - N_{ref5} \\ &= TKN_4 - 0.015 \, COD_5 - N_{ref4} \qquad (8\text{-}29) \end{aligned}$$

$$COD_4 = S_c \, COD_0 \qquad (8\text{-}30)$$

Nitrates formed = Influent TKN − TKN synthesized − TKN in effluent
TKN synthesized = 0.015 x biodegradable COD

Anoxic Zone   Aerobic Zone
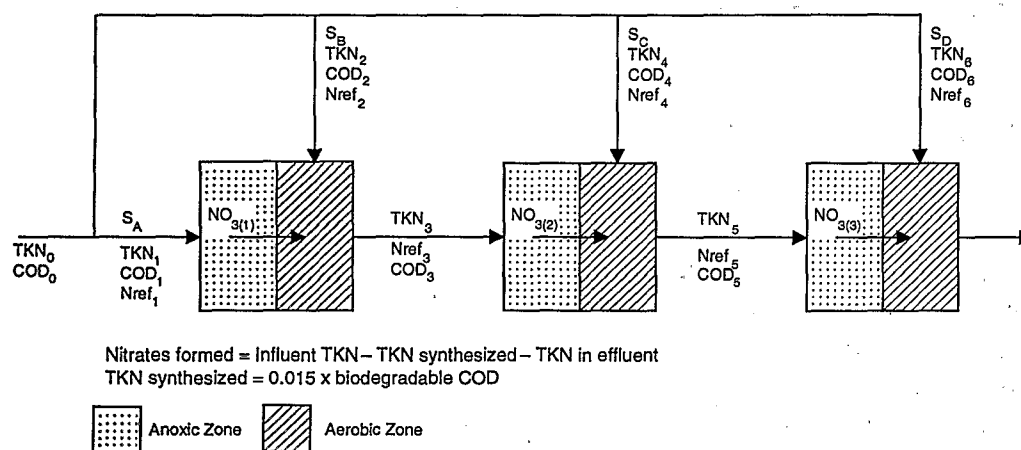
Figure 8-16. Multi-anoxic zone step-feed process.

$$COD_5 = COD_4 - 2.9 \, NO_{3(2)} \qquad (8\text{-}31)$$

$$COD_6 = S_D \, COD_o \qquad (8\text{-}32)$$

$$TKN_3 = TKN_2 + N_{ref1} \qquad (8\text{-}33)$$

$$TKN_4 = S_C \, TKN_o \qquad (8\text{-}34)$$

$$TKN_5 = TKN_4 + N_{ref3} \qquad (8\text{-}35)$$

$$TKN_6 = S_D \, TKN_o \qquad (8\text{-}36)$$

$$N_{ref3} = N_{ref1} + N_{ref2} \qquad (8\text{-}37)$$

$$N_{ref5} = N_{ref3} + N_{ref4} \qquad (8\text{-}38)$$

If adequate COD and retention time are available to denitrify all nitrate, the effluent will only contain TKN introduced with the flow S_D and N_{ref5}. These equations can be modified to account for nitrates or ammonia in the return sludge flow or to make further refinements in the nitrogen balance due to biomass synthesis in each of the reactor zones. A computation sheet containing the preceding equations can be used to optimize the step-feed ratios and percent removals. The theoretical removals as a function of the influent COD:TKN are presented in Figure 8-17.

Figure 8-17 demonstrates step-feed processes can theoretically achieve >90-percent 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 post aeration.

8.2.3.4 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
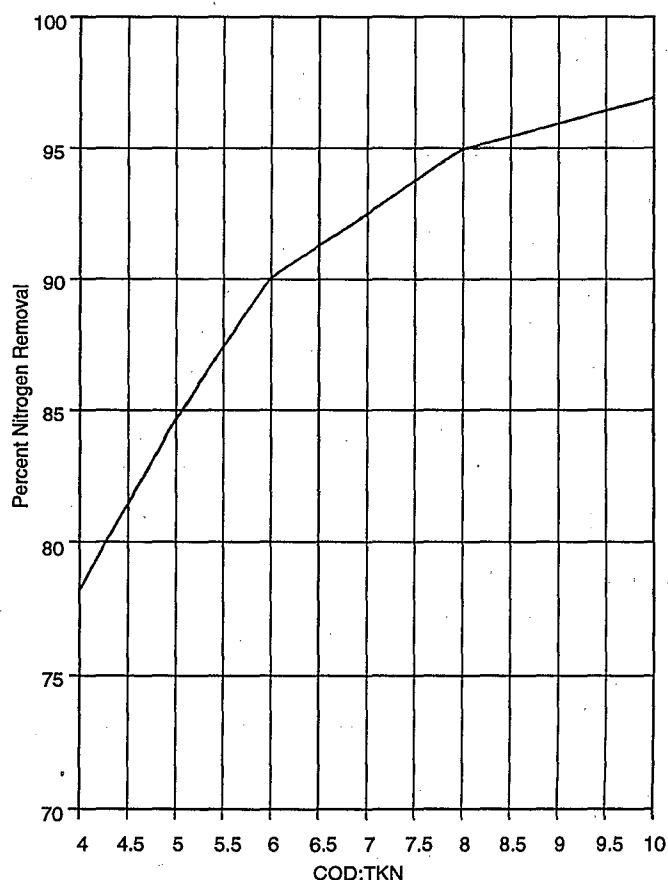

Figure 8-17. Theoretical percent nitrogen removal as a function of COD:TKN for a triple anoxic zone process with step feed.

266

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 microorganisms. 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 (9). 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 8-6.

8.2.4 Multi-anoxic Zones

8.2.4.1 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

Table 8-6. Monitoring Requirements and Rationale for Bardenpho Reactors (9)

| | 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

with conversion to a proprietary nitrogen removal process, if it is applicable.

One study of aeration cycling was conducted at the Blue Plains Pilot Plant in Washington, DC (14). Although the performance during the study was subject to upsets, 84 percent nitrogen removal was achieved. The F/M ratio (BOD_5/MLVSS/d) was kept at 0.1, which was low enough to permit a mixed culture for nitrification and denitrification. Researchers achieved TN concentrations <7 mg/L at the onset of the study in cold weather conditions.

Process upsets at Blue Plains were attributed to experimentation with unsuccessful modifications that were later discontinued and to the lack of experience associated with the first attempt at the CNR technology. The process was susceptible to sludge bulking, necessitating operational changes that reduced denitrification. The Blue Plains facility had experienced bulking under normal operation; therefore, the cycling of the aerators could not be stated with certainty to be the causative factor. The denitrification process was interrupted by ceasing the aeration cycling to eliminate the bulking organisms. The addition of $FeCl_3$ to primary influent was initiated in warm weather to promote phosphorus removal, but $FeCl_3$ also reduced the COD:TKN from 10:1 to 8:1 causing a decrease in denitrification. Despite these operational modifications and upsets, the process did achieve TN effluent concentrations <8 mg/L and greater than 80-percent nitrogen removal.

Denitrification was observed to be the limiting rate reaction in summer months at Blue Plains. The nitrification rate was not observed to be sensitive to pulsing aeration and was observed to be relatively constant during the aeration cycles. Denitrification rates were observed to decrease, ostensibly as a result of depletion of the carbon source. This effect could be surmounted by providing step feeding during the anoxic cycle. The implication from these studies to the design of cyclic operation confirms the prudence of performing nitrification and denitrification rate tests to determine cycling frequency and duration. It was further suggested that long θ_c's were necessary to ensure a high ratio of nitrifiers to denitrifiers to optimize CNR performance (12).

TN removals of 80 percent in summer and just under 80 percent in winter were achieved at the Owego, New York, wastewater treatment plant (15). 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 Figure 8-18. Subsequent investigations (16) 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 (Figure 8-19). The
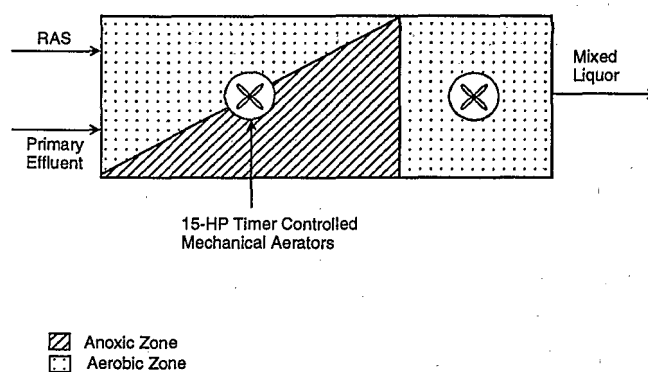

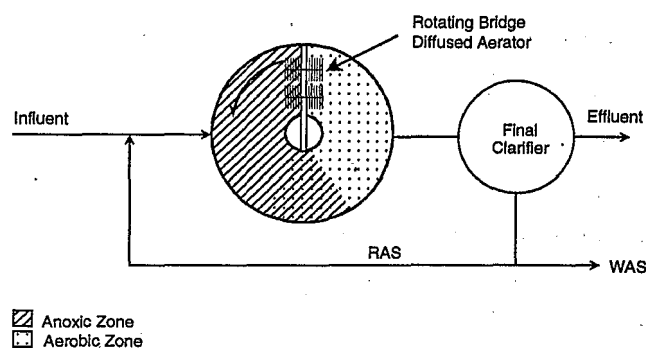Figure 8-18. Town of Owego, NY, water pollution control

Figure 8-19. Schreiber process.

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 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.

8.2.4.2 Typical Design Criteria

CNR design incorporates similar considerations as single anoxic zone processes. Aeration capacity, solids retention time (SRT), 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 (15); calculations should determine the

adequacy of existing reactor basin volume, aeration capacity, and settling capacity. Design criteria are presented in Table 8-7.

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.

8.2.4.3 Process Performance

CNR systems can consistently produce effluent TN concentrations <8 mg/L and >80 percent TN removal. Al-

Table 8-7. Cyclical Aeration Design Criteria

Parameter	CNR at Owego	Schreiber
F/M, g/BOD$_5$/g MLVSS/d	0.06–0.13	0.05
Aerator on, min	15–45	*
Cycle off, min	15–30	*
θ_c, d	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	2,600–4,000	2,000–7,000

* Load oriented with turbidity control.

though the CNR process has not been used or investigated extensively, pilot and full-scale operating results are presented in Table 8-8, 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.

8.2.4.4 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 (15). The recommendation for a minimum of three basins in series is predicated on the provision of step feeding to the downstream basins. The CNR process with step feed is analogous to the Miyaji process (7) discussed in Section 8.2.3. The process performance considerations introduced in Section 8.2.3.3 indicated that process performance is enhanced by increasing the number of reactors.

Existing plants that use a plug flow configuration can be modified to achieve CNR by dividing the reactor into com-

Table 8-8. Cyclical Aeration Operating Results

Process	CNR (16)	CNR (15)	CNR (12)	Schreiber (17)	Schreiber (17)
Location	Barnstable, MA	Owego, NY	Blue Plains Wash., DC	Clayton County, GA	Jackson, TN
Q, m^3/d	5,450	1,820	N/A	8,970	31,260
HRT, hr	9	13–16	10.1	N/A	N/A
θ_c, d	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/d	0.08[d] 0.24[e]	0.089	0.089		

[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
N/A = Data not available

269

partments with baffles. The provision of step-feed capability will distribute the BOD load to each compartment and ensure an exogenous carbon source for denitrification during each aerator off-cycle. Step-feed inputs should be located one-half to three-quarters of the distance from the influent end of the tank. The organic carbon demand in the downstream portions of the tank will vary primarily with temperature. Rapid depletion of available carbon has been observed in the upstream portion of the aeration tank in summer months when warmer temperatures cause increased growth rates. Though not always necessary to meet discharge requirements in a CNR process, higher return sludge pumping rates can improve performance. Internal recycle pumping of mixed liquor from the effluent to the influent end of the aeration tank, or from the point of the last anoxic cycle zone to the head end of the aeration tank may also improve performance. In many retrofit situations where the plant is operating below design capacity, the existing return sludge pumps may provide sufficient recycle flow (i.e., 100–200 percent of plant influent) to meet effluent TN concentrations <8 mg/L. Step feeding may not be required in cold weather as BOD depletion may not occur as rapidly in the head end of the tank.

Aeration cycling can be accomplished at timed intervals or by using set point signals from DO probes. Multiple aerators in a single basin can have staggered operation to avoid excessive power current draws. Timers should have programmable control to permit aeration cycling corresponding to diurnal and workday/weekend variations. Systems that use mechanical surface aerators may not require submerged mixers during the anoxic phase. The residual turbulence imparted by mechanical surface aerators has been observed to suspend the floc to provide sufficient liquid-solids contact for a 30-minute anoxic cycle (15). Although solids separation may not be of concern with 15–45 minute anoxic cycles and higher MLSS, low-energy mixing may enhance substrate availability for the biomass. Cycling of air in diffuser systems to different compartments of the basin can be accomplished by using electrically operated butterfly valves. Oxygen transfer in excess of that required for complete nitrification during the aerobic cycle should be avoided; at existing installations this will require careful monitoring to determine the actual oxygen transfer capacity of the aeration system. Excess aeration (DO >2 mg/L) will delay denitrification by prolonging the lag period required for DO depletion. Subsurface mixers may be desirable in diffuser systems since the residual turbulence with diffused air is less than that provided by mechanical surface aerators.

Several unique concepts are incorporated in the design of the Schreiber process (Figure 8-19). These include the mixed liquor velocity, rotating bridge velocity, DO profile (a transitional curve dependent on bridge location), and the variation of DO concentration for subsequent bridge

revolutions. Since the rate of oxygen transfer demand varies, automated systems are provided with the Schreiber process. Types of programmable control that can be provided include DO probes to cycle blowers on and off or a patented O_2 minimizer. The O_2 minimizer is designed to vary aeration in response to the turbidity of a settled MLSS sample, which varies with the quantity of oxygen supplied to satisfy biochemical and nitrogenous oxygen demand.

8.2.5 Oxidation Ditches

8.2.5.1 Background and Process Description

Oxidation ditch processes include technologies that use looped trenches that provide a continuous circulation path for the wastewater. Aerators within the flow path simultaneously provide the necessary aeration of the waste stream as well as fluid propulsion. This type of aeration results in an aerobic zone for nitrification in that portion of the loop immediately downstream of the aerators, while immediately upstream of the aerators, oxygen deficient conditions usually prevail, thereby providing anoxic conditions for denitrification.

Oxidation ditch technology was pioneered in the late 1950s by the Dutch engineer Pasveer, whose purpose was to develop an inexpensive wastewater treatment process for small communities that required a minimum of operator attention. The Pasveer ditch consisted of an oval loop, in which flow was induced and aerated by a rotating aerator. Over 3,000 ditches were installed based on the Pasveer configuration. The foremost limitation of Pasveer's process was the areal requirement to accommodate the reactor loop, particularly for large populations. Subsequent studies in the late 1960s and early 1970s surmounted these limitations by modifying the reactor configurations.

The next significant oxidation ditch investigation after the Pasveer effort was initiated in Austria (18). Treatment improvement was attempted at the Vienna-Blumenthal, Austria, plant by using multiple aerators within each loop and then connecting the tandem loops in series instead of in parallel. The process schematic is illustrated in Figure 8-20. Denitrification in the second loop was attempted by reducing the aeration within the loop, thereby providing a process configuration similar to the postdenitrification Wuhrmann concept. Ironically, nitrification-denitrification was observed to occur between the rotors of the first tank, which receives the raw influent, because of higher respiration rates afforded from the higher COD load. Subsequent studies using oxidation ditches hypothesized that denitrification could also be occurring at the interior portions of the floc particle where an oxygen deficient condition existed (1,19).
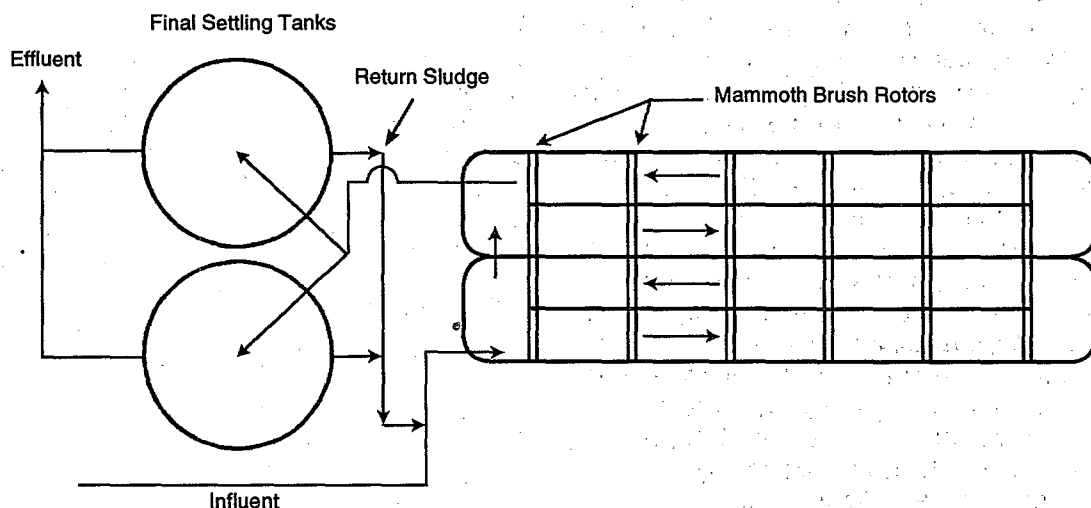
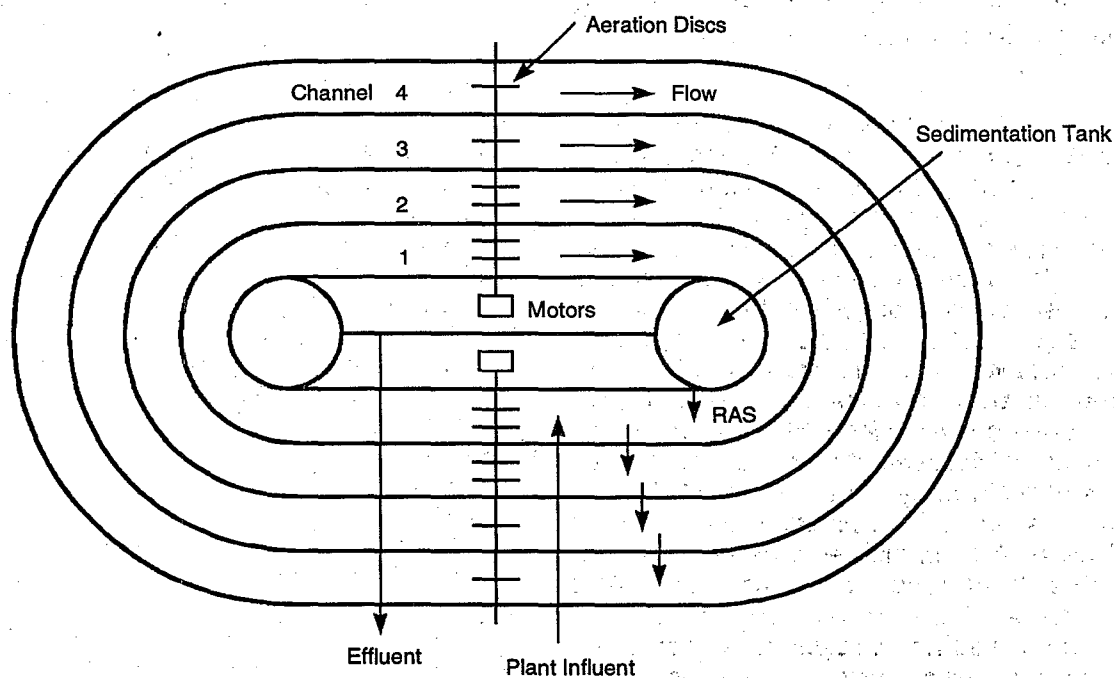**Figure 8-20. Vienna-Blumenthal Wastewater Treatment Plant.**

Figure 8-21. Orbal oxidation ditch.

The areal requirement of the early oxidation ditch designs was reduced with the development of larger, more efficient aeration devices, and the development of deeper basins. In addition, the flow path per unit area was increased by either arranging the flow loops concentrically or by folding the flow oval in half, as shown in Figures 8-21 and 8-22, respectively. The former configuration became trademarked as the Orbal Process, and the latter the Carrousel System by their developers. The Orbal patent by Envirex applies only to the aeration disk equipment itself and not to the reactor configuration process flow train. The

license for the Carrousel System is held by Eimco Process Equipment Company of Salt Lake City, Utah. The capitalized form of the words will be used in this manual to distinguish the trademarked systems.

The earliest reference to the Carrousel process was made in 1968 (20), when it was held as an innovative solution to implementing ditch technology to serve large populations. The Carrousel process developers cited the requirement for excessive surface area (and thus space), an infeasible capital expenditure for numerous surface
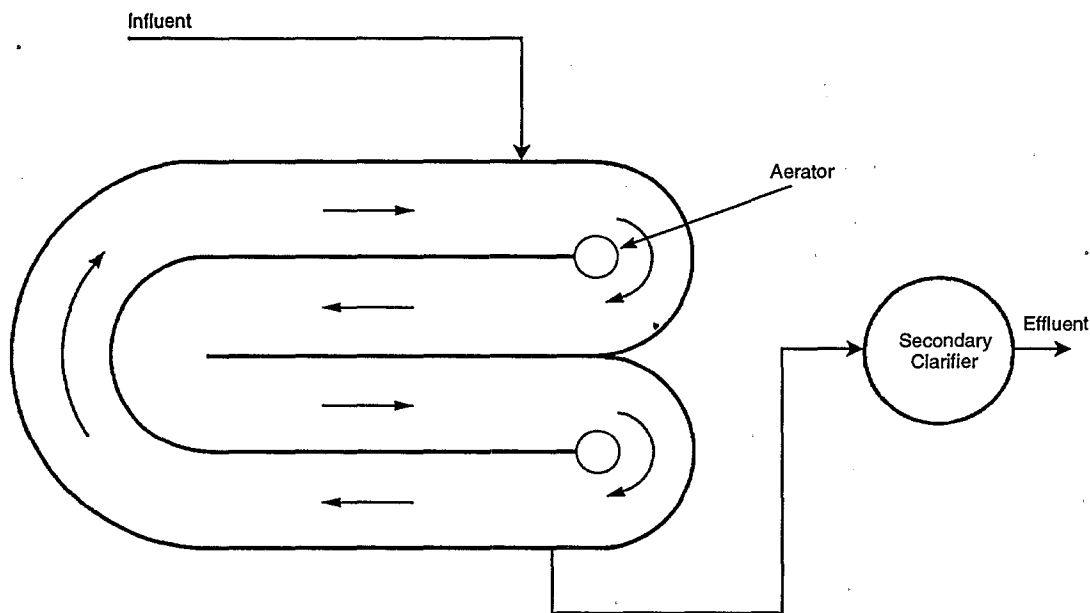
271

Figure 8-22. Carrousel oxidation ditch.

aerators, and the concomitant energy costs. The Carrousel design was developed to provide adequate aeration and to maintain adequate flow velocities, while simultaneously maximizing utilization of surface area. Licensed by EIMCO since 1976, the Carrousel process is a proprietary process in the United States which has been installed at numerous U.S. facilities and at over 500 sites worldwide. The Carrousel process is widely used in Holland (its place of origin) and to a lesser extent in Germany and England.

The orbal process was developed primarily in South Africa as another optional arrangement of oxidation ditch channels that would maximize utility of land area and minimize costs of aeration equipment and power, while simultaneously providing alternating aerobic and anoxic conditions. The orbal system consists of concentrically arranged channels, as shown in Figure 8-21. Raw wastewater or primary effluent is introduced into the innermost or outermost channel and is conveyed to the other channels via ports at the bottom of the channel dividing walls. The installation date of the first orbal plant was not available from a review of the literature, but 50 orbal plants were in operation by 1972 in South Africa (21).

The nitrification-denitrification option of the patented Orbal process by Envirex is termed the Sim-Pre process, schematically illustrated in Figure 8-23. Sim-Pre is an acronym for simultaneous nitrification-denitrification/pre-denitrification. The Sim-Pre process incorporates an internal recycle from the innermost to the outermost channel. This aspect of the process represents predenitrification. The simultaneous nitrification-denitrification phase occurs in the first aeration channel. Since the aera-

tion demand exceeds the supply, anoxic conditions are attained along the flow path in the area upstream of the aerators. Thus, the outer channel of the Sim-Pre process is operated analogous to a conventional oxidation ditch.

I. Kruger of Denmark, represented in the United States by I. Kruger, Inc., in conjunction with the University of Denmark, has developed two ditch-type processes for nitrogen removal. These employ multiple ditches of large total volume and flexible, simple timing controls of rotors (mixing and aeration) and weir levels. The patented BioDenitro process is designed to remove nitrogen, while the BioDenipho is a modification to remove phosphorus and nitrogen. The basic BioDenitro configuration consists of two identical aeration ditch tanks and a clarifier. The process can be applied to any type of complete mix aeration tank system, as long as both aeration (e.g., brush aerators or diffused air) and mixing equipment (typical slow-speed propeller mixer) are installed. The separate provision of mixing and aeration thus allows tanks or ditches with high-sidewall depths, e.g., 4.6 m (15 ft).

However, instead of creating anoxic and aerobic zones within each tank as in a conventional oxidation ditch, the BioDenitro process achieves aerobic or anoxic conditions alternately within each looped reactor. The wastewater feed is alternated between the two tanks to provide a carbon donor source for the desired microbial reactions. Two designs of the BioDenitro plant are available. The type DE plant consists of two oxidation ditches and a final clarifier. The operating sequence is shown in Figure 8-24. This four-phase alternating influent contact operation is an application of a process first introduced by Christensen (3). The BioDenipho process for phosphorus removal can
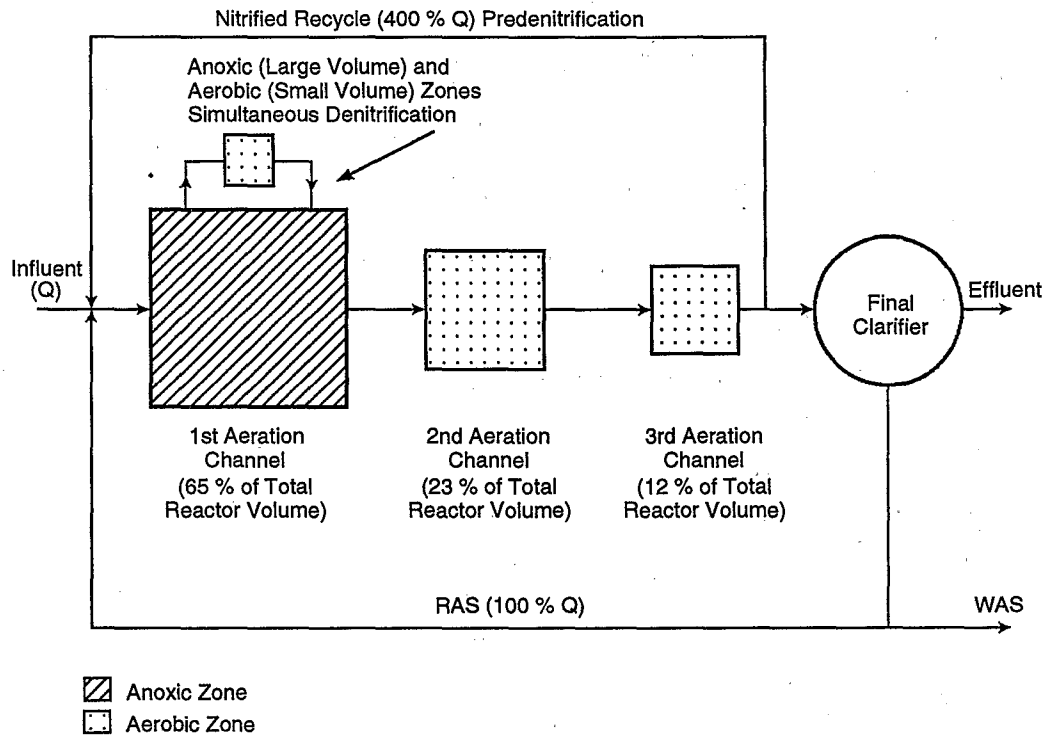
Figure 8-23.   Orbal Sim-Pre process.

Phases A & C (Main Operational Phases): 60-90 min.
Phases B & D (Intermediate Operational Phases): 15-30 min.

⊥⊥⊥⊥   Rotor Mixing   N = Nitrification
╫╫╫   Aeration   DN = Denitrification (Anoxic)
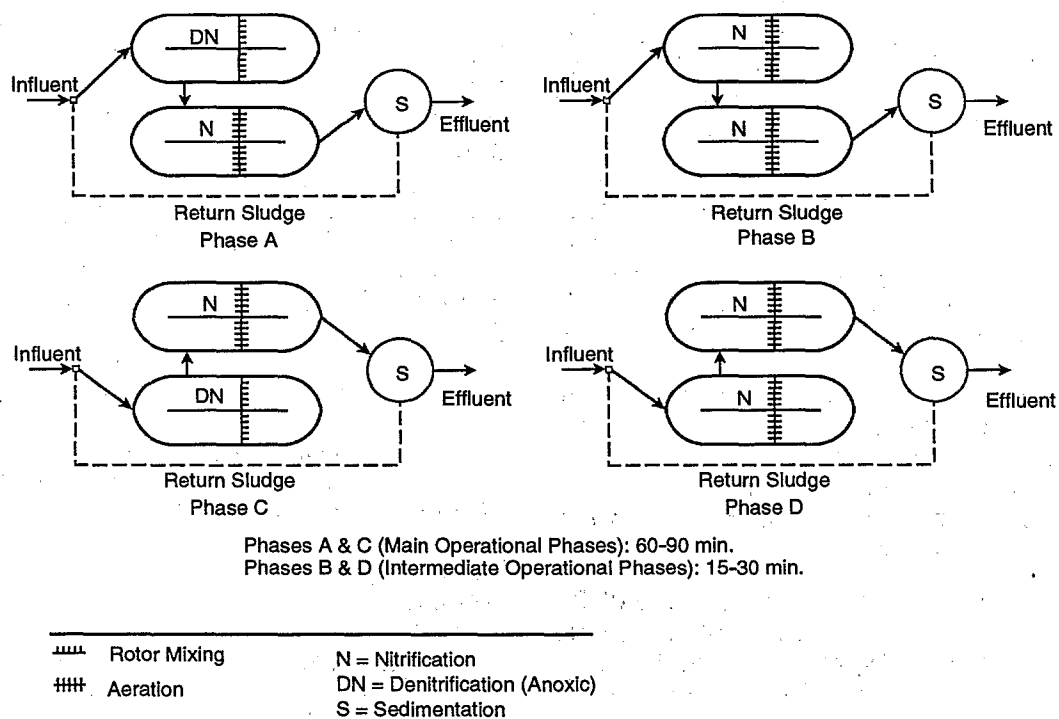   S = Sedimentation

Figure 8-24.   Kruger BioDenitro process (Type DE).

273

be provided by including an anaerobic phase to the BioDenitro process.

The second BioDenitro process design uses the T-ditch configuration in which first and third oxidation ditches primarily provide denitrification and settling; the middle oxidation ditch always serves as an aeration and flow distribution unit. The BioDenitro T process is a six-phase cycle and is schematically illustrated in Figure 8-25. This system is very much a sequencing batch reactor of unique design and operating features. As shown in Figures 8-24 and 8-25, neither the T nor the DE process uses an internal recycle. The T process does not use an RAS recycle since separate clarifiers are not provided.

8.2.5.2 Typical Design Criteria

Typical design criteria are unavailable for oxidation ditches since their designs vary. However, criteria used from existing plants are presented in Table 8-9.

Design for denitrification in oxidation ditches is similar to other anoxic reactor designs. Methods of computing the nitrogen available for nitrification and denitrification were

introduced in Section 8.2.2; these are applicable for oxidation ditch designs as well.

Oxidation ditch technologies are generally operated in the extended aeration mode, with long hydraulic and solids retention times and higher MLSS than are commonly used in activated sludge plants. Conceptually, the oxidation ditch configuration is an endless channel. Only a portion of the mixed liquor is withdrawn in each cycle, providing a high internal recycle ratio. The orbal configuration incorporates properties similar to plug flow, depending on the transfer and distribution method of mixed liquor between successive channels.

The size of the ditch should be determined based on maintaining channel velocities of 0.3–0.6 m/s (1–2 fps), or a loop flow time of 10–45 min. The design of the channels can vary, but will incorporate considerations used in the design of single anoxic zone systems, such as sludge concentrations, F/M ratios, RAS rate, temperature extremes, and desired effluent quality. The F/M ratio has been suggested as the critical design parameter (21),
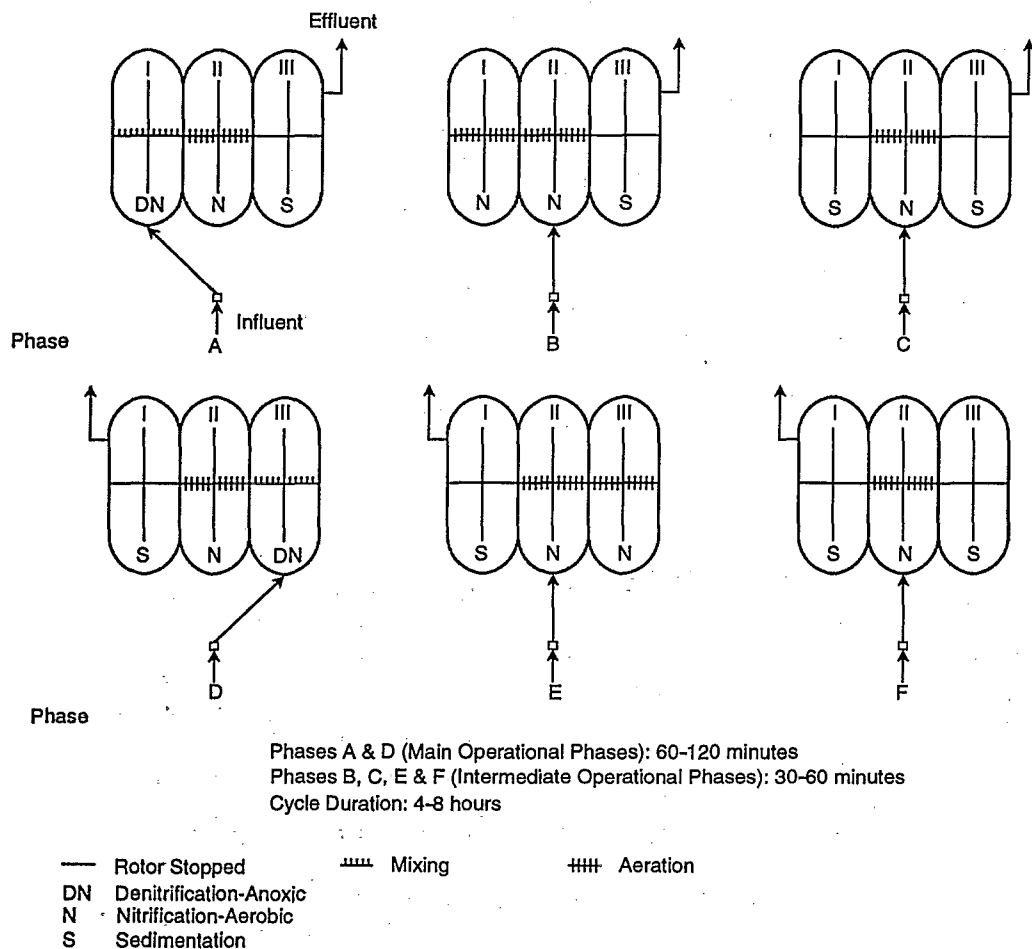

Phases A & D (Main Operational Phases): 60-120 minutes
Phases B, C, E & F (Intermediate Operational Phases): 30-60 minutes
Cycle Duration: 4-8 hours

——— Rotor Stopped   ⊔⊔⊔⊔ Mixing   ++++ Aeration
DN   Denitrification-Anoxic
N · Nitrification-Aerobic
S   Sedimentation

Figure 8-25. Kruger BioDenitro process (Type T).

274

Table 8-9. Operating Parameters at Various Oxidation Ditches

Type	Oxidation Ditch (22)	Orbal (23)	Orbal (21)	Orbal (21)	Orbal (1) Extended Aeration	Orbal (1) Split Feed	Oxidation Ditch (19)	BioDenitro DE	Bio-Denitro T
Location	Vienna-Blumenthal, Austria		Modder-fontein, S. Africa	S. Wit-bank, S. Africa	Huntsville, Texas	Huntsville, Texas	Carrol-wood, Florida	Frederiks-sund, Denmark	Odense, Denmark
Q, m^3/d	41,290	189	2,385	167	98	114	13,250	6,000	15,000
MLSS, mg/L	5,800	4,300	3,030	8,830	8,000	8,800	3,060	3,000–5,500[e]	3,000[e]
F/M, g BOD_5/g MLSS/d	0.17	0.22[a]	0.093	0.03	0.027[b]	0.015[b]	NA	0.08[e]	
RAS, %Q	190	200	115		110	95	—	30–80[e]	NA
θ_c, d	7		15		>50	>50	44	15–30[e]	15–30[e]
HRT, hr	7		11	31	33	28	17	14	22
Volumetric loading, g $BOD/m^3/d$	985	144–240	416	194–226	120	90	150[c]	707	282
lb BOD/1,000 cu ft/d	61	9–15	26	15–18	8	6	9[d]	44	18

[a] COD/MLSS   [b] g BOD_5/g MLVSS/d   [c] g $COD/m^3/d$ [d] lb COD/1,000 cu ft/d   [e] Design value

and can be used as a design basis if the minimum sludge concentration is known.

The required aeration capacity is difficult to ascertain accurately during design. Factors that affect aeration requirements include wastewater characteristics, flow variations, temperature, and RAS rate. Further design considerations, such as sizing aerobic and anaerobic or anoxic zones for TN removal, are complex and difficult to predict. However, a suggested approach would be to design for complete nitrification between the aerators. By varying the oxygen transfer and monitoring the DO profile in the channel, the operating nitrification rate can be ascertained for the actual COD and temperature of the wastewater. The oxygen transfer rate can then be further adjusted to achieve oxygen deficient conditions downstream of the aerators. Several researchers have suggested that nitrification and denitrification can best be obtained by specifying the required anoxic volume and then monitoring and controlling the DO profiles (24,25). Also, flow and load equalization has been suggested as a means to reduce shock loads, or in the case of Orbal configurations, enhanced denitrification has been reported by step-feeding influent to the inner channels (1).

The use of temporal sequencing in the design of the Kruger BioDenitro process is unique. As shown in Figures 8-24 and 8-25, this method of operation allows operator flexibility to optimize the anoxic-aerobic volumes by varying the time duration of the specific phases. During anoxic phase, the aeration is stopped and the sludge is maintained in suspension by submerged propeller mixers.

Thus no oxygen is introduced while the anoxic zone is created. Similarly, the aeration equipment is operated to establish a defined aerobic volume. Typically, the standard BioDenitro type DE process volume is proportioned to 57–67 percent for nitrification and 33–43 percent for denitrification. The denitrification volume can be changed to accommodate fluctuations in flow, level, or temperature simply by changing the duration of the cycle. Further flexibility may be obtained by adding the third aerobic ditch of the T process. The third ditch ensures that all mixed liquor is aerated before entering the final clarifiers and permits the main ditches to increase the portion of time allotted for denitrification.

Design of an Orbal Sim-Pre process follows the standard procedure used by Envirex for the design of its Orbal process. Denitrification is afforded by the internal recycling of mixed liquor from the innermost to the outermost channel. The Orbal process design by Envirex uses a single factor of 0.93 to account for α, β, temperature, and elevation. Each of the three channels are designed to achieve different DO levels. The suppliers of the Sim-Pre process recommend that the channels should be designed for 0, 1, and 2 mg/L DO for the outer, middle, and inner channels, respectively, although the actual DO will vary widely along the channel. The channel volume split in the Sim-Pre mode is generally 65/23/12 for outer, middle, and inner channels, corresponding to a base oxygen demand split of 70/18/12, respectively. Thus, for a total O_2 requirement of 908 kg (2,000 lb) O_2/d, the corresponding oxygen delivery to each channel would be 590, 209,

and 109 kg (1,300, 460, and 240 lb) O_2/d. With these requirements, the number of discs, immersion, and rotational speed can be selected.

Typical design criteria for the Orbal process are presented in Table 8-10.

8.2.5.3 Process Performance

Oxidation ditch designs vary and, consequently, oxidation ditch performance also varies. Performance results from a number of facilities are presented in Table 8-11. Nitrogen removals range from 65 to 97 percent. The performance results reported by one manufacturer (26) indicate that removals as high as 90 percent have been attained.

In certain cases, Envirex has guaranteed process effluent total nitrogen of 3 mg/L.

The Kruger process incorporates a unique feature of flexibility due to its timed alternating aeration and mixing sequences, incorporating aspects of an SBR. However, it differs from many SBRs in that the Kruger process provides continuous influent and effluent flows. The phase lengths may be varied to achieve specific treatment objectives, or to respond to variations in load, flow, or temperature. Kruger claims that the operator can vary the operation of the plant within two minutes in response to an impetus from the control room. The process can be provided with a sophisticated monitoring and operation

Table 8-10. Design Parameters for Orbal and Orbal Sim-Pre Process

Flow, m³/d	BOD Load, g/m³	MLSS, mg/L	θ_c, d	HRT, hr	Depth, m
<760	200	4,000–5,000	31–38	24	1.2–2.4
760–1,889	240	4,000–5,000	26–32	20	1.5–2.4
1,890–3,784	240–288	4,000–5,000	20–32	16.6–20	1.8–3.0
3,785–7,569	288	4,000–5,000	21–27	16.6	2.4–3.7
>7,570	320	5,000–6,000	24–29	15	2.4–3.7

Table 8-11. Nitrogen Removal Performance for Various Oxidation Ditch-Type Plants

Type	Oxidation Ditch (22)	Orbal (21)	Orbal (21)	Orbal Step Feed (1)	Oxidation Ditch (19)	Oxidation Ditch (19)	Kruger T (27)	Kruger DE (27)
Location	Vienna-Blumenthal, Austria	South Witbank, South Africa	Modderfontein, South Africa	Huntsville, TX	Carrolwood, FL	Frankfort, KY	Faaborg, Denmark	Frederikssund, Denmark
FLOW								
m³/d (mgd)	41,256 (10.9)	227 (0.06)	2,271 (0.6)	2,271 (0.6)	13,248 (3.5)	14,383 (3.8)	15,000 (3.96)	8,327 (2.2)
BOD (COD), mg/L								
Influent	245	319	200	168	(250)	(205)		300
Effluent	12	3	5	6	(20)	(23)	6	9
TKN, mg/L								
Influent	30	52.3	34	19.4	25	16.8		36
Effluent	3.1	8.4	10	0.7	0.6	1.1		
BOD (COD):TKN								
Influent		6.1	5.9	8.7	(10.0)	(12.2)		8.3
NH₄-N, mg/L								
Influent	17.9	39.2	21	17.8				
Effluent	3.6	6.7	7.3	0.6	0.3		4.8	0.5
NOₓ-N, mg/L								
Effluent	0.9	0.7	2.2	1.1	0.04	2.5	4.3	1.5
Total N, mg/L								
Effluent	4	9.1	12.2	1.8	0.64	3.6	9.1	3.5
% N removal	87	86	65	91	97	79	80	90

PC-based control system. For example, the plant can have, in the PLC or timer control, preprogrammed/preset operational modes for the weekend, weekdays, summer, high-load, etc. The operator would have the discretion to activate these operating sequences.

Oxidation ditch technologies were developed to accentuate simplicity, minimize operator control, and provide cost and energy savings. The ditch configuration is the most energy efficient design available because it conserves fluid momentum. This flow regime augments stability, consistency, and energy efficiency (27). The key operational parameters in an oxidation ditch are the oxygen transfer rate and MLSS. The MLSS may be controlled by the waste activated sludge (WAS) rates from the clarifier (or from the ditch for Kruger T). In a single oxidation ditch loop, aeration can be optimized by placing the aerators at selected points along the flow path. Orbal configurations can incorporate features of a compartmented plug flow reactor by maintaining individual loops at a desired average DO level. For example, the outermost channel can be maintained at aerobic conditions, while the second channel is maintained at anoxic conditions. Nitrates formed in the aerobic channel will pass to the second (anoxic) channel, where denitrification can occur. Step-feeding raw wastewater or influent to the anoxic channel will ensure a carbon source to promote substrate nitrogen respiration. If active recycling of mixed liquor between channels is not provided, the oxygen transfer capacity is the most direct means to control nitrate loads. Orbal plants can reduce aeration equipment costs by extending a single aerator shaft across all the channels. Oxygen transfer efficiency and capacity may be adjusted in Kruger, Orbal, and conventional single-loop oxidation ditches by adjusting the submergence of the aeration discs or rotors, varying the rotational speed, or changing the number of aerators on each shaft in the channel. Cycling of the aerators can also be used, as long as sufficient alternate impetus is provided to propel the mixed liquor in the channel.

The Orbal configuration also is conducive to split-feed options and has brought about enhanced nitrogen removals. The following hypotheses were offered to explain the efficacy of split-feed approaches (1):

1. Higher DO levels and enhanced nitrification were achieved by lowering the organic load and oxygen demand in the first channel.

2. Raw wastewater addition into the second channel lowered the oxygen concentration in the channel and extended the denitrification zone.

3. Raw wastewater addition to the second channel provided an exogenous carbon source to promote denitrification.

8.2.5.4 Process Design Features

Oxidation ditches are a hybrid flow regime and incorporate aspects of plug flow and a CSTR. The concentric channels of Orbal configurations have been mathematically proven to approach an ideal plug flow reactor (21,28). As the wastewater recycle (i.e., the average number of passes around the loop) is increased, the system was also determined to approach an ideal CSTR. At long retention times (extended aeration), the limiting nutrient concentration is lower in a plug flow reactor. Although superior effluent quality may be realized as a result of the advantages of a plug flow regime, the provision of equalization facilities for plug flow regimes has been suggested because of their lesser ability to accommodate shock loads.

As discussed, alternating oxygen levels can be provided by varying the aeration capacity or by installing concentric channels in the Orbal configuration and maintaining the desired condition within each channel. The Orbal arrangement can be further modified to promote denitrification by providing step-feed capability. Additional influent feed points may also be included in a single oxidation ditch loop in an attempt to ensure an exogenous carbon source.

Several investigators have commented on the susceptibility of oxidation ditch technologies to poor settling characteristics (23,29) which were postulated to be exacerbated by excessive rotational energy input by the aeration mechanism, causing floc shear. Poor settling characteristics were also attributed to placement of an anoxic zone near the effluent withdrawal for the clarifiers. Presumably, the settling was hindered by entrainment of nitrogen gas in the floc, causing a floating sludge. This condition was corrected by ensuring that the ditch effluent withdrawal point was preceded by an aerated zone.

Aeration of oxidation ditches is also a critical operational parameter. Continuous monitoring of the DO is required to develop an oxygen profile that provides anoxic and aerobic zones. The oxidation ditch can be automated by using DO probes to coordinate aerator cycles, weir motors, or a combination of the two. To a certain extent, ditches are somewhat self-regulating in the case of hydraulic load increases. As the water surface of the ditch rises, the rotors will be more submerged, thereby imparting a higher oxygen transfer to the waste. This feature can be modulated mechanically if desired.

8.2.6 Sequencing Batch Reactors

8.2.6.1 Background and Process Description

An SBR, as it is commonly referred to today, is a fill-and-draw, variable reactor volume technology. The prototype for the activated sludge concept (30) was developed on a fill-and-draw basis. Shortly after that initial study, the

emphasis switched to continuous flow "conventional" activated sludge. Further developments with SBR technology were not pursued because of limitations of equipment and engineering experience—namely the ability to dissipate the hydraulic energy of the effluent discharge, operator attention, and expertise, and the susceptibility of clogging of the air diffusers during the settling period. Recent innovations in aeration devices, control logic, level sensors, solenoids, and hydraulic energy dissipators have surmounted these limitations and revitalized interest in SBR technology. The resurgence of interest in SBRs was initially limited to small treatment applications; however, the need for greater treatment efficiencies due to increasingly stringent effluent limits has resulted in the adoption of SBR technology in installations as large as 660 L/s (15 mgd) (31).

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. *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 ef-

ficiencies 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. Sludge wastage will typically be performed during the idle phase.

A typical SBR process sequence schematic is shown in Figure 8-26.

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 suf-
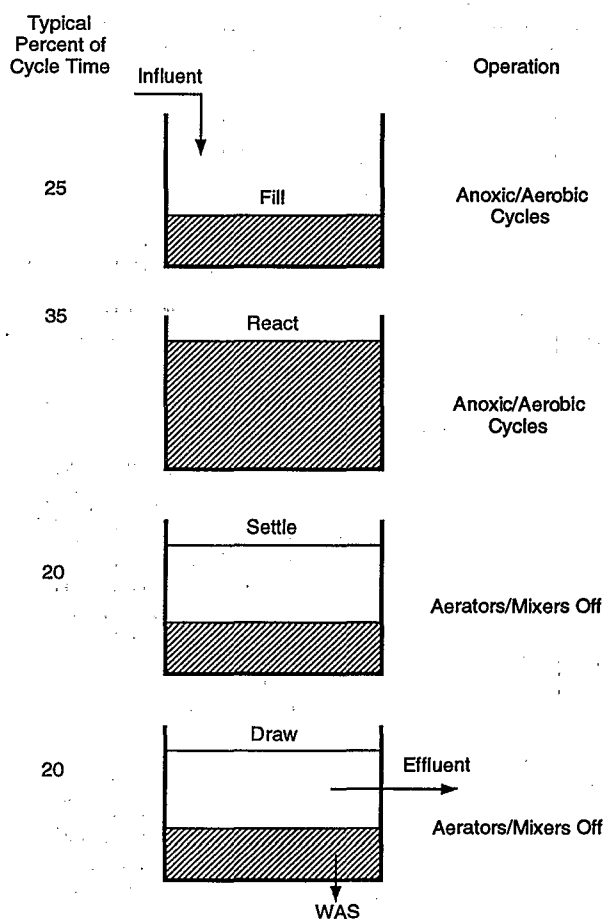

Figure 8-26. Sequencing batch reactor.

278

ficient tankage exists to equalize or accommodate the additional influent.

Several proprietary process and equipment innovations have been developed to enhance treatment, simplify operation, or control sludge characteristics. 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.

- *Aqua SBR*—The Aqua SBR system provided by Aqua-Aerobic Systems, Inc., 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 SVIs.

- *Omniflo*—Jet Tech, Inc., 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.

- *Fluidyne*—The Fluidyne Corp. offers a system with effluent decanters fixed in position to the reactor wall. The device excludes MLSS entry during aeration. These systems also commonly employ jet aeration with a combination of aeration and static conditions during fill.

- *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 (32). A process schematic is presented in Figure 8-27.
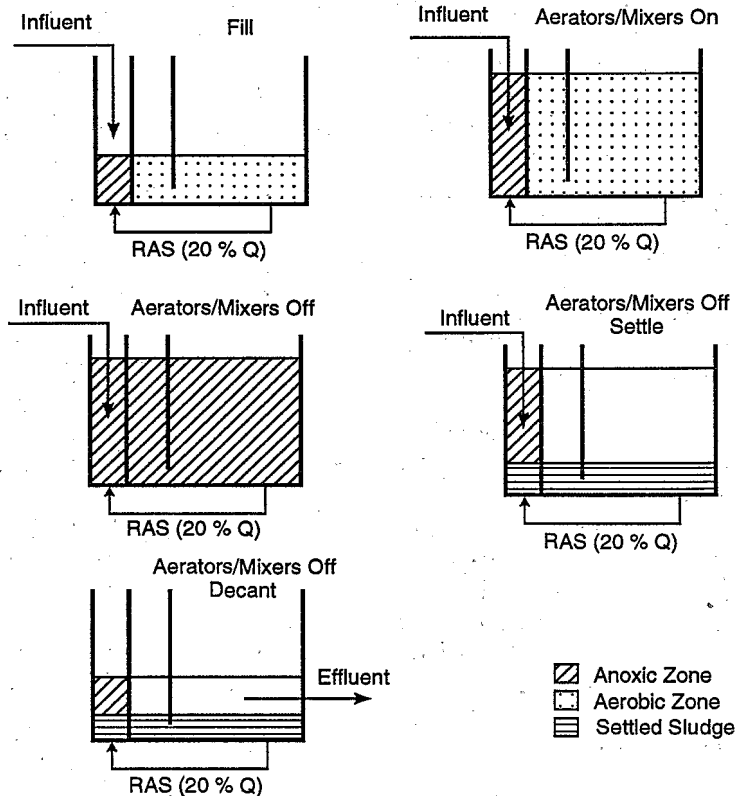

Figure 8-27. Cyclical Activated Sludge System.

• *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 Figure 8-28. The distinguishing features 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. Since influent is received during all phases, ICEAS does not offer total quiescence during the settle phase, 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 microorganism, 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.

As noted in the previous section, some Kruger ditch configurations (Figure 8-25) are variations of the SBR concept in a unique looped reactor design.

8.2.6.2 Typical Design Criteria

A unified approach to SBR technology has yet to be developed (33); 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 which will determine whether one reactor will suffice (generally for flows <2 L/s [0.05 mgd]) or whether a 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 3–10 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 four-hour duration; two hours allocated to fill/aeration/anoxic react, one hour to settling, and one 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 full-scale plants have varied from 2 to 24 hours (34). A suggested strategy is presented in Figure 8-29. Some typical design criteria are presented in Table 8-12.

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 3–4 m (10–13 ft) and design top water levels are typically 4.3–5.5 m (14–18 ft). A freeboard of 1 m (3 ft) is common. 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 a settled mixed liquor and is similar to that in a conventional clarifier underflow (35). 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.
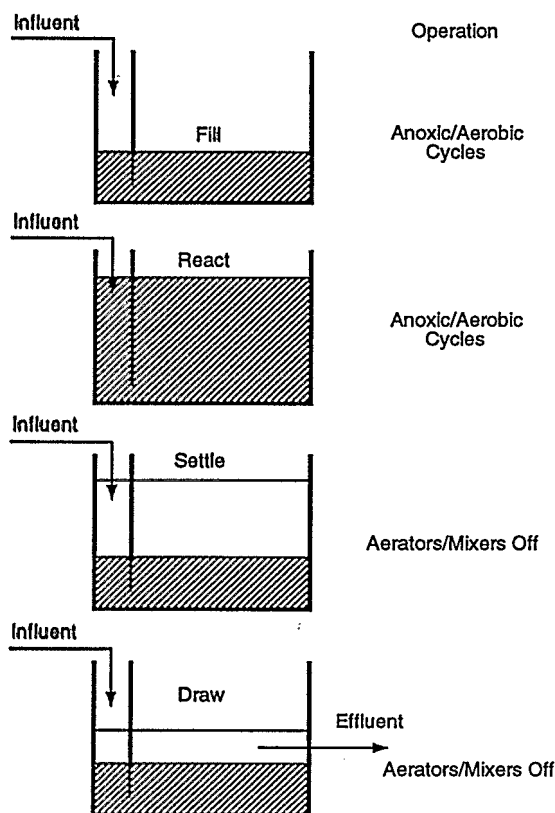

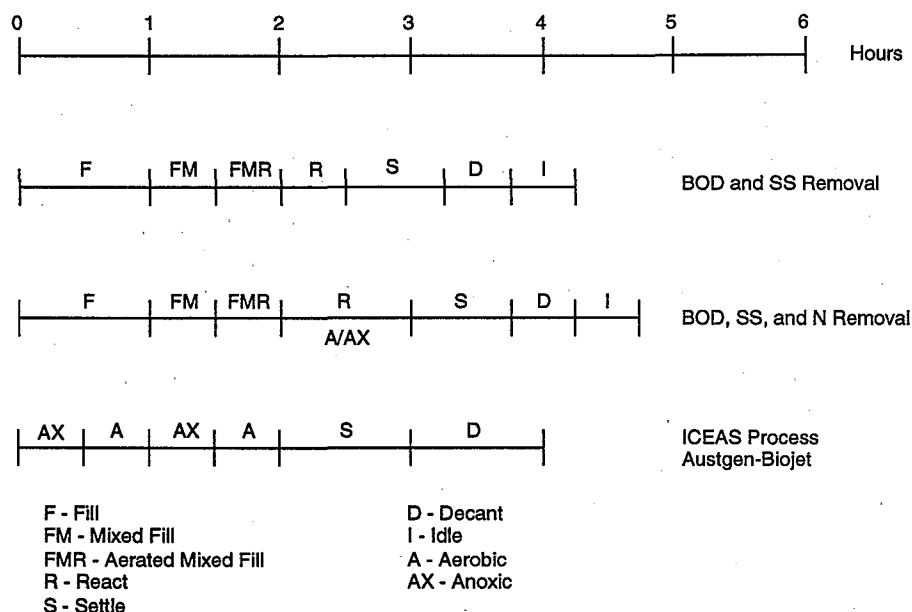**Figure 8-28. Intermittent Cycle Extended Aeration System.**

Figure 8-29. Suggested operating strategies for SBR systems.

Table 8-12. Typical Design Criteria for Sequencing Batch Reactors

Parameter	SBR	ICEAS
BOD load, g/d/m³	80–240	
Cycle time, hr		
Fill (aeration)	1–3	
Settle	0.7–1	
Draw	0.5–1.5	
MLSS, mg/L	2,300–5,000	
MLVSS, mg/L	1,500–3,500	
HRT, hr	15–40	36–50
θ_c, d	20–40	—
F/M, g BOD/g MLVSS/d	0.05–0.20	0.04–0.06

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.

8.2.6.3 Process Performance

Nitrogen removal in SBR systems can be considerably higher than in conventional activated sludge systems. Performance results from full-scale facilities are provided in Table 8-13. Available data on full-scale SBR facilities detailing nitrogen removal are limited because of the relatively recent application of this technology.

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:

- The SBR system can tolerate shock loads and peak flows because of the equalizing basin characteristics of the fill phase.

- Periodic effluent discharge may permit retention of reactor contents until desired clarity or treatment quality is achieved.

- 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.

- No RAS or internal recycles are required; however, some systems (e.g., CASS) include recycle to an antecedent basin or selector chamber.

- 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.

- Filamentous growth can be controlled by operational strategies along with adjustments during the fill phase.

SBR operation is somewhat more sophisticated than other systems of comparable size. However, the advent of reliable automation of liquid level sensors with programmable logic controllers and decanting devices significantly simplifies operation. Particular attention to process performance and monitoring is required to optimize the performance and to determine the optimum aeration cycle frequency.

281

Table 8-13. Summary of SBR Plant Operating Data (36)

Plant	Flow m³/d, mgd	Influent BOD₅, mg/L	Influent TKN (Total N), mg/L	Effluent TKN (Total N), mg/L	Influent NH₄⁺-N, mg/L	Effluent NH₄⁺-N, mg/L	Effluent NOₓ-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*	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*	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*	84
Jet Tech Oak Pt., MI	227 (0.06)	142	N/A	N/A	19.0	0.6	2.8	3.4*	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*	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

N/A - Data not available
* Based on effluent NH₄⁺-N + Nₒₓ-N

8.2.6.4 Process Design Features

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 sludge characteristics. Several criteria have been proposed that can be used to design an appropriate selector (8,37,38). The CASS process by Transenviro is a proprietary SBR that includes an integral selector as part of the process.

The process control of an SBR requires relatively sophisticated coordination of probes, valves, timers, and level sensors. 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.

Maintenance of the desired solids inventory is facilitated in an SBR since aeration and settling occur in the same reactor and the entire sludge mass may be retained if required. There is no set interval dedicated to sludge wasting in an SBR, but wasting is only conducted as performance requirements dictate.

The mixing requirements of an SBR reactor are similar to flow-through systems. Since all reactions occur in the same basin, some aeration systems cannot generally be used to provide mixing during anoxic cycles. However, the jet aeration systems offer independent control of mix-

ing and oxygen transfer. 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. Mixing considerations in SBR processes during anoxic cycles are similar to considerations discussed for single and oval anoxic zone processes.

8.3 Process Selection Considerations

Selection of the most cost-effective and efficient type of single-sludge system for application to a particular wastewater treatment facility will be influenced by a number of factors, including:

- Effluent limits

- Wastewater characteristics

- Site constraints

- Existing facilities

- Cost

The following discussion of selection factors will focus on single-sludge systems but will include some consideration of separate-sludge systems. In many cases, combining at least BOD removal and nitrification into a single-sludge system will be cost effective considering the additional cost associated with separating these two processes. However, there will be situations where effluent limits, wastewater characteristics, or physical constraints require a separate-stage denitrification system.

8.3.1 Effluent Limits

The first criteria for selecting process alternatives will be the permitted discharge limits. In general, process selection will be driven by the form and degree of nitrogen removal required with requirements for phosphorus removal being secondary. Treatment alternatives may be broadly divided into systems that remove ammonia (nitrification), those that remove nitrogen (nitrification and denitrification), and those that remove both nitrogen and phosphorus.

8.3.1.1 Nitrification

Permits limiting ammonia, and not TN, will require a treatment system consisting of an aerobic section only, using a single-sludge or a separate-sludge system. However, even if not required by the discharge permit, it may be desirable to include denitrification in the treatment system. Denitrification, which reduces the nitrates produced during the nitrification process, offers features in a single-sludge system that can reduce operating costs. These include the elimination or the reduction of chemical required for pH adjustment in low-alkalinity wastewaters

subject to pH depression as a result of the nitrification process and the reduction of aeration requirements because of the consumption of wastewater carbon (BOD) by nitrates. There is a potential advantage to including denitrification in a system that includes biological phosphorus removal. Nitrates interfere with biological phosphorus removal. Denitrification will reduce or eliminate the recycle of nitrates back to the anaerobic zone and thus improve the system's phosphorus removal capabilities. The presence of nitrates can also create problems in the final settling tanks. If the DO level in the influent to the settling tank is low, denitrification will occur when a source of wastewater carbon is available or more likely because of endogenous metabolism. These conditions are likely to occur below the sludge blanket. If denitrification is significant, nitrogen gas bubbles will become attached to sludge particles causing them to rise. An operational solution to this problem is to maintain the minimum possible sludge blanket in the final clarifiers. Rapid sludge removal collectors may also be utilized.

8.3.1.2 Denitrification

Permits that limit the TN level will require the treatment process to nitrify and denitrify. As discussed above, single-sludge systems are capable of complete nitrification. Process selection becomes more a matter of how to implement denitrification. The options available, in order of increasing complexity, are single-sludge systems with alternating aerobic/anoxic phases, single-sludge systems with one aerobic/anoxic zone, single-sludge systems with two or more aerobic/anoxic zones, and finally, separate-sludge systems.

These basic options encompass a variety of treatment processes—some that require relatively simple operational modifications to implement and others that require more extensive additions to the existing facility.

8.3.1.3 Treatment Performance

The following discussion of nitrogen removal performance of the various treatment alternatives is intended as a general guideline for the selection of alternatives for further evaluation. The performance attributed to each process is generally conservative. Actual treatment plant performance data for various processes have been provided elsewhere in the manual. In some cases, however, there is little full-scale operating data from which to make engineering judgements. The discussion of performance is divided into ranges of effluent TN. In reality, process performance can not be so neatly delineated because treatment efficiency depends on many factors that make generalizations difficult. These factors include wastewater characteristics, variability of flows and loads, temperature, presence of inhibiting compounds, and the design features of the plant (e.g., flexibility of operation, clarifier loadings, and recycle flows).

In many cases it may be possible to improve the performance of a process by modifying its operation or by adding relatively minor equipment or process improvements. For example, the nitrogen removal performance of a system with a single anoxic zone may be improved—within limits—by increasing the mixed liquor recycle rate. A CNR system's performance may be improved by adding step feed or an internal recycle. Methanol trimming may be added to a second anoxic zone to improve denitrification kinetics. If there is adequate excess capacity, a second anoxic zone may be added.

When evaluating wastewater treatment alternatives for biological nutrient removal (BNR), the tendency to lock the design criteria for a particular process into a narrow set of numbers should be avoided. For example, the Bardenpho process is generally a low-load process with relatively long hydraulic retention times. If justified by the specific circumstances, however, it can be designed to operate at a higher rate, similar to the A^2/O or VIP process. The effluent limits and wastewater characteristics should be the driving force in the design and not preconceived limits on design criteria.

Actual plant operating data for the various processes may be found in Section 8.2, which also presents detailed descriptions and design criteria of the single-sludge processes. The following discussion approaches the task of selecting a treatment process for nitrogen removal from the standpoint of the maximum allowed permit level for the discharge of TN. Where appropriate, the phosphorus removal capability of the system is also included.

a. Total Nitrogen: 8–12 mg/L

Basically, any of the single anoxic zone systems will be capable of achieving TN residuals of 8–12 mg/L in a typical domestic wastewater on an average annual basis. Multi-anoxic zone systems such as some types of oxidation ditches or the CNR process would also be candidates for providing cost-effective treatment.

An MLE type of approach (Figure 8-3) can be relatively easily retrofitted into an existing wastewater treatment plant through the installation of baffles, mixers, and an internal recycle capacity. In a diffused air system, the diffusers would have to be relocated to create an anoxic zone and to ensure that there is adequate air for nitrification in the oxic zone. Some diffusers could be left in the anoxic zone to provide mixing if the DO concentration were kept at a low level. An MLE process, which includes an internal recycle, is capable of meeting an 8-mg/L TN permit. Without the internal recycle, the process becomes the basic Ludzack-Ettinger schematic (Figure 8-2) which should be able to meet a 12-mg/L TN permit.

The UCT, A^2/O, and VIP processes may be implemented for a TN limit in the 8-12 mg/L range. However, these processes also include an anaerobic zone or selector for

phosphorus removal and bulking control. Selection of one of these processes will depend more on the wastewater characteristics. The UCT or VIP process would be favored when a high degree of phosphorus removal is required and when the $TBOD_5$:TP is low (<20:1). Both the A^2/O and VIP processes are high-rate systems with relatively low θ_c's and short hydraulic retention times. As such, they may offer more opportunities for implementation in an existing treatment plant with limited excess capacity.

Oxidation ditches are capable of performing in this range of TN removal. The performance of a ditch system cannot be predicted with as much certainty as other activated sludge systems. Acceptable performance will require field monitoring for optimizing the operation, such as determining the DO profile. The process can then be optimized by adjusting the DO level, which is done by turning aerators on or off or by controlling the oxygen transfer rate by varying the aerator submergence and/or horsepower.

SBRs can produce an effluent with <8 mg/L TN but close attention to operating conditions is required to optimize performance. SBRs are well suited for relatively small but highly variable flows, since equalization of flows is inherent in the process.

The CNR process has been demonstrated to produce a TN effluent of <10 mg/L and, with attention to operating conditions, can consistently achieve TN levels of ≤8 mg/L. This may require the ability to step feed with multiple compartments, especially in warmer weather, to provide a carbon source for denitrification. In existing plants with little excess capacity, CNR offers the potential to meet a seasonal TN permit with relatively small modifications. During the warmer periods, when reaction rates for carbonaceous BOD and nitrification are higher, the aerobic volume of the basin may be smaller, thus allowing capacity for the anoxic operating condition. During colder periods, if all the existing tank volume is required for the aerobic processes, the system would not be able to denitrify. If the permit was not based on a seasonal TN limit, then either additional tank volume would be required or a separate-stage denitrification system would be needed. If the permit places only a seasonal restriction on TN, then the separate-stage denitrification system would not be required.

b. Total Nitrogen: 6–8 mg/L

Typically, an effluent TN limit of 8 mg/L is presented as the limit of performance for single anoxic zone systems, such as for the A^2/O, UCT, and VIP processes. However, these systems should be capable of producing effluent TN concentrations of 6–8 mg/L with further enhancements. As the TN limit decreases, the internal mixed liquor recycle rate from the aerobic to the anoxic zone must increase. The ability to recycle 100–400 percent of the flow should be provided. Also, the design criteria must be

more conservative in order to provide the treatment efficiency and reliability required. A second anoxic zone may be added with methanol addition to reduce remaining nitrates.

Dual anoxic zone processes should be considered in this range of performance although they are capable of even greater TN removal. The Bardenpho system, for example, is capable of achieving TN removals to <3 mg/L. However, in the 6–8 mg/L TN range, it is possible to implement the Bardenpho process without effluent filtration if adequate settling capacity is provided.

Oxidation ditches are capable of achieving effluent TN levels down to 6 mg/L. The literature reports a wide variety of treatment efficiencies for ditch systems principally because the operating conditions and wastewater characteristics vary so greatly. The inherent internal recycle rate of a ditch system often exceeds 100 times the influent flow. This tends to dilute the influent BOD, which improves the competitive advantage of the nitrifying organisms. The nitrifiers are able to compete for the available DO over a larger fraction of the aerobic volume. Thus, nitrification is maintained over a larger fraction of the mixed liquor volume. However, the dilution of BOD, which is the carbon source for denitrification, also tends to decrease the denitrification rate. Another key to reliable nitrogen removal performance in an oxidation ditch is to provide the ability to closely monitor and control the DO, as was discussed in Section 8.2.5.4.

The BioDenitro process by Kruger can consistently produce an effluent with a TN level of 6 mg/L. Effluent concentrations of 3 mg/L have been obtained, but consistent nitrogen removal to this effluent concentration has not yet been demonstrated.

c. Total Nitrogen: 3–6 mg/L

Dual anoxic zone processes must be considered to achieve TN residuals of 3–6 mg/L because of the additional denitrification that is required and provided by the second anoxic zone. The Bardenpho process is typically the single-sludge system applied at TN limits ≤3 mg/L. This process was developed as a nitrogen removal process in contrast to the A/O or A²/O processes, which typically operate at higher loading rates and shorter hydraulic retention times (HRTs).

The Modified UCT process has two anoxic zones, but the second anoxic zone follows in series after the first. As was discussed in detail in Section 8.2.3, the Modified UCT performance is similar to single anoxic zone processes. Single anoxic zone processes are not able to achieve effluent nitrogen concentrations of 3–6 mg/L for typical wastewaters. Since the Modified UCT is intended as a compromise between nitrogen and phosphorus removal, this process would be expected to be similar or slightly better to that of the UCT process.

Separate-sludge systems can also be designed for TN permit levels of ≤3 mg/L. Separate-stage denitrification using a downflow filter may be a desirable approach if a strict TSS limit (<10 mg/L) is also imposed by the discharge permit. If effluent filtration is not desired as a component of the denitrification system, then an upflow packed-bed reactor may be used for denitrification. During the warmer months, it may be possible to implement cyclical aeration in the nitrification stage to achieve denitrification and reduce the methanol requirements in the denitrification reactor, thus saving operating costs.

8.3.1.4 Phosphorus Removal

Many discharge permits limiting the discharge of TN will also include a TP limit. Very low levels of effluent TP can be achieved through biological processes (for a detailed discussion of these systems, see Reference 39).

Where moderate levels of nitrogen removal are required, single-sludge, single anoxic zone systems such as the A²/O, VIP, and UCT processes can be applied. These systems can produce effluents with <1 mg/L TP although results are variable. Selection among the A²/O, UCT, and VIP processes depends primarily on the wastewater characteristics. Reliable performance to ≤1 mg/L will require chemical precipitation in addition to biological removal. As discussed previously, dual anoxic zone systems are typically applied where low levels of TN are also required. In this case, the modified or five-stage Bardenpho process would be considered since it includes an anaerobic zone at the beginning of the process train. Effluent TP levels of <3 mg/L have been reported with these systems.

To achieve a ≤1-mg/L TP limit, and to consistently produce an effluent with <2 mg/L TP, provisions for chemical addition should be included. TP levels below 0.5 mg/L may require effluent filtration in addition to chemical treatment. A phosphate detergent ban or use of a selector may allow plants to achieve effluent phosphorus concentrations consistently below 1 mg/L; however, testing would be required to verify this.

Biological nutrient removal systems designed for both nitrogen and phosphorus removal present some unique operating considerations since the two processes have conflicting requirements. The mechanism for biological phosphorus removal in a mainstream process such as UCT, A²/O, VIP, or modified Bardenpho is through the assimilation of enhanced levels of phosphorus in the bacteria selected by the anaerobic zone and subsequent elimination of this phosphorus through waste sludge. Therefore, biological phosphorus removal is optimized at a shorter θ_c than is typically employed for nitrification. Sludge production and subsequent phosphorus removal is promoted at the shorter θ_c's. Generally, the compromise is to operate at the shortest θ_c required to achieve effluent nitrogen limits.

Nitrates also present a problem for biological phosphorus removal systems. For "luxury" uptake of phosphorus to proceed, the bacteria must first assimilate organic substrates in the anaerobic zone and store them within their cells. Nitrates that may be present in the RAS allow facultative heterotrophic bacteria to compete for soluble substrates required by the phosphorus-removing bacteria. This can degrade the phosphorus-removal performance of systems such as the A^2/O or modified Bardenpho, which recycle RAS directly to the anaerobic zone. The UCT and VIP processes avoid this problem by returning the RAS to the anoxic zone where the nitrates are reduced by denitrification. Mixed liquor is then recycled from downstream of the anoxic zone to the anaerobic zone.

A key selection criteria, therefore, for BNR systems which must remove both nitrogen and phosphorus is the $TBOD_5$:TP ratio. If this ratio is greater than 20:1, then the recycle of nitrates to the anaerobic zone may not be a problem because there is an adequate supply of excess organic substrate. In this case, either the A^2/O or the modified Bardenpho process should be considered for stricter limits on TN. These systems do not have the costs associated with an additional recycle from the anoxic to the anaerobic zone. If the $TBOD_5$:TP is less than 20:1, then the VIP or UCT process should be considered. For the same reasons, the TKN:BOD is important. A high TKN:BOD may indicate there is insufficient carbon to denitrify completely, resulting in a carryover of nitrates to the anaerobic zone.

For wastewater treatment plants that must remove phosphorus and are required to nitrify but not denitrify, it may still be advantageous to include denitrification as a component of a single-sludge system. This will reduce the interference of nitrates with the biological phosphorus removal mechanism and will also provide the cost savings associated with the return of alkalinity and reduction in aeration requirements.

Mainstream biological phosphorus-removal systems will require effluent filtration in order to achieve low levels of TP (<0.5 mg/L). This is because the solids that are carried by the clarified effluent will contain an enhanced level of phosphorus, which can have a significant impact on the effluent TP levels. Typical activated sludge solids will contain approximately 2 percent phosphorus, whereas the solids from a biological phosphorus removal system will contain typically 4–8 percent phosphorus.

8.3.1.5 Time Basis of Permit Limits

A permit with a TN limit that is based on a yearly average obviously provides greater flexibility than a permit that requires the same discharge limit on the basis of a maximum monthly average. Similarly, a monthly average permit provides more freedom than one based on a maximum weekly average. With a shorter time period, fewer excursions above the permit limit can be tolerated.

The shorter time periods will therefore require a more conservative process design approach to improve reliability. Operating experience at biological nutrient removal facilities has indicated that typically the maximum month effluent TN will be 1.4 times the average annual value (40). Phosphorus removal is more variable. Typically, the maximum month effluent TP will be 2 times the average annual value.

8.3.1.6 Seasonal Limits

A permit requiring nitrification or nitrogen removal only in the summer offers more opportunities for process selection than does a permit that requires year-round nitrogen control. For example, a plant that has limited excess capacity may not be able to implement denitrification using CNR year-round, but may be able to do so in the summer when reaction rates are greater. Because the aerobic volume required for nitrification is less in the summer, there is generally more flexibility in existing plants to achieve the seasonal permit limits through operational modifications of the existing process. Seasonal permits are common for ammonia removal when the concern is related to fish toxicity. Having to nitrify only in the summer reduces the aeration tank volume required and thus reduces capital costs.

Most instances of noncompliance with effluent nitrogen limits occur in the first month when the seasonal restrictions apply. To ensure that the system can achieve the level of removal when required, the wastewater temperature during the month prior to the month the seasonal limits are to take effect should be included when establishing the design temperature.

8.3.1.7 Solids Limits

A well-operated treatment plant with conservatively designed final clarifiers should be able to consistently produce an effluent with <15 mg/L TSS. Effluent <10 mg/L will generally require effluent filtration.

Once filtration is required, more consideration may be given to separate-stage nitrification-denitrification because the sand filter with methanol addition can also provide the required denitrification. However, the analysis must also consider the present worth of the added operations and maintenance costs. Since separate-stage denitrification requires methanol addition, the savings in aeration costs and chemical for alkalinity adjustment associated with single-sludge systems will not be realized.

A low level of TN required by a discharge permit does not necessarily imply that filtration must be added. Although effluent solids do contain organic nitrogen, the amount of organic nitrogen that would appear in the TN of the final effluent may be small. A TSS level of 30 mg/L in the final effluent would contribute only approximately 2 mg/L to the TN level. A requirement for very low levels of

TN (<1 mg/L) will require filtration because of the particulate organic nitrogen and the nondegradable soluble organic nitrogen fraction. Often discharge permits requiring very low levels of effluent TN also limit effluent TSS to a level where filtration is required.

8.3.2 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, with a high ratio favoring denitrification. Also significant is the presence of readily available organic carbon, which is indicated by $SBOD_5$:BOD_5. A high proportion of soluble and readily degradable BOD would favor denitrification and improve reaction rates.

Wastewater temperature is a critical parameter since 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.

The pH of a wastewater also affects nitrification and denitrification rates. The optimum pH range for nitrification is generally accepted to be 6.5–8.5. For denitrification the optimum pH is 7.0–8.0. Since 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 percent 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 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. 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 increase (Figure 1-5).

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.

8.3.3 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. 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. This process, which was developed to optimize biological phosphorus removal, operates at a total θ_c of 5–10 d under warm weather conditions compared to the modified Bardenpho or UCT processes, which typically operate at θ_c's of 10–25 d. 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 systems offer the possibility of a high degree of treatment in a relatively small space. Since 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.

8.3.4 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.

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 plug flow 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 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 (Section 6.6). 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.

8.3.5 Costs

8.3.5.1 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, 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.

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 15-20 percent more than a single-sludge system. General guidelines for costs of various components of a biological nutrient removal system have been published (40). 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.

8.3.5.2 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 percent. 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 cost 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 utilize 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.

8.4 Design Considerations

8.4.1 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. 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: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 de-

nitrification. A BOD: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.

8.4.2 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 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.

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.

DO 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 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 1–2 mg/L.

8.4.3 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, or variable-speed, variable-depth controls on the Kruger ditches.

8.4.4 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 volumes as large as four times the plant influent flow may be required. Larger pumping volumes are impractical since the marginal increase in nitrogen removal via internal recycle decreases significantly for recycle rates >400 percent of the influent flow. This concept is discussed in detail in Section 8.2.3.3. 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.

8.4.5 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.

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.

8.4.6 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 due 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 DO. 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, since 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 (38).

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 prior to 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 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.

8.4.7 Selectors

Several researchers have observed poor sludge settling characteristics in nitrogen removal processes (14,38,41). Nuisance filamentous organisms *Microthrix parvicella, Sphaerotilus natans, Nocardia,* and Types 021N and 1701 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 microorganism. Low F/M organisms exhibit a higher growth rate at low substrate levels (38). 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 outcompete filamentous organisms. Thus, bulking sludges due 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 (41), low BOD:N and BOD:P (42), 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 (43). Mixing return sludge with influent wastewater in one or more in-series contact chambers for a short duration prior to directing the stream into a complete mix basin has been suggested (44–46); the pre-react chamber described was termed a "selector" since 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 (41). This observation has been confirmed at several full-scale plants (8).

The following have been recommended for effective selector design (43):

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 percent of soluble substrate.

3. Microbial activity (determined from the substrate uptake rate) should be maintained as high as possible.

While aerobic, anaerobic, and anoxic selectors have all been found to control bulking effectively, the type of aeration was determined to influence selector performance (8). 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 due to the plug flow regime or the anaerobic conditions, or both, the flow regime and anaerobic condition in concert, could not be determined. While both aerobic and anaerobic selectors can control bulking, anaerobic selectors can also provide the benefit of phosphorus removal without requiring additional aeration capacity.

8.5 Process Design Examples

8.5.1 Introduction

The process design examples in this section illustrate three of the different process types. The design examples illustrate sizing calculations for two different plant scenarios (Figures 2-5 and 2-6) with two different sets of effluent limitations (Table 2-10). Both plants in the scenarios from previous examples are activated sludge plants and have an average daily design flow of 220 L/s (5 mgd). Plant A does not have primary settling or separate digestion. Plant B 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 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 TP is also considered for that design example. Table 8-14 summarizes the two sets of effluent limits.

The process design examples in this chapter expand upon the examples in Chapter 6, which illustrated the design of single-sludge nitrification systems and sizing of the aeration tanks. The following examples illustrate process design and sizing for single-sludge nitrification-denitrification systems, which primarily involves sizing of the

Table 8-14. Design Example Effluent Limits

		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

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.

The following design features are some of the factors to be considered during the 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:TKN is adequate (at least 3:1) such that wastewater carbon is the carbon source. Internal recycle pumps should generally be sized to provide an upper limit of 3 to 4 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.

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:

- Design Example No. 1. Single Anoxic Zone—Plant B (complex plant) and less stringent limits.

- Design Example No. 2. Dual Anoxic Zones—Plant B (complex plant) with more stringent limits.

- Design Example No. 3. Multiple Anoxic Phases (cyclical)—Plant A (simple plant) with less stringent limits.

- Design Example No. 4. Dual Anoxic Zones—Plant A (simple plant) with more stringent limits.

8.5.2 Summary of Design Procedures

The following is an outline of procedures used in designing single-sludge nitrification-denitrification systems:

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 as shown in Tables 2-15 and 2-16 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 re-

quired 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 prior to 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, since the denitrification rate is dependent on the anoxic F/M ratio (availability of COD as the carbon source).

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.

7. Size the post-aeration zone to achieve a residual DO level of 1–2 mg/L prior to the secondary clarifier.

8. Determine WAS and RAS pumping requirements to cover the full range of possible conditions.

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.

8.5.3 Design Example No. 1: Plant B with Less Stringent Limits

From the mass balances in Table 2-16, the inlet wastewater characteristics (primary effluent with recycles) and secondary effluent characteristics for the more complex plant without nitrification-denitrification are as follows:

	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 minimum DO = 2.0 mg/L
Reactor MLSS = 3,000 mg/L	Secondary effluent NH$_4^+$-N = 1.0 mg/L
MLVSS = 63%	Secondary effluent Org-N = 2.0 mg/L
Reactor pH range = 7.0–7.6	Secondary effluent NO$_3^-$-N = 5.0 mg/L

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-percent TN removal efficiency are required. A single anoxic zone-type process, such as the MLE process, is adequate to meet these limits. The plant configuration is as shown in Figure 2-6, 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 percent) will be affected by the solids retention time (θ_c) and that the recycle stream characteristics will remain as shown in the mass balance in Table 2-16. Therefore, the mass balance is unchanged.

Removal Requirements Across Secondary Process:

At average annual loadings (concentrations in mg/L equivalents)

BOD$_5$ Removed = (18,925 m^3/d) (97 − 3 mg/L)/1,000 = 1,779 kg (3,922 lb)/d

TN Removed = (18,925 m^3/d) (29.5 − 8.0 mg/L)/1,000 = 407 kg (897 lb)/d

TN Removed in Waste Solids by Assimilation = (18,925 m^3/d) (3.0 mg/L)/1,000

$$= 57 \text{ kg (125 lb)/d}$$

TN Removed by Nitrification-Denitrification = 407 − 57 = 350 kg (772 lb)/d

294

8.5.3 Design Example No. 1 (continued)

1. *Sizing of nitrification zone.* Similar to the design procedure outlined in Section 6.4.2.2, determine the size of the nitrification reactor.

Calculate maximum nitrifier growth rate, $\hat{\mu}_N$:

$$\hat{\mu}_N = 0.47 \ e^{0.098(T-15)} \tag{3-14}$$

$\hat{\mu}_N = 0.47 \ d^{-1}$ at 15°C

Calculate specific growth rate, μ_N:

$$\mu_N = \hat{\mu}_N \ [N/(K_N + N)] \tag{3-10}$$

where:

> $N = 1.0$ mg/L
> $K_N = 1.0$ mg/L
> $\mu_N = (0.47)[1.0/(1.0 + 1.0)] = 0.23 \ d^{-1}$

Calculate the minimum solids retention time, θ_c^m:

$$\theta_c^m = 1/\mu_N \tag{3-13}$$

$\theta_c^m = 1/0.23$

$\theta_c^m = 4.35$ d

Calculate design solids retention time, θ_c^d, using a design factor (Chapter 6):

$$\theta_c^d = (PF) \ (SF) \ (\theta_c^m)$$

where:

> $PF = 1.56$ (Table 2.18)
> $SF = 1.25$
> $\theta_c^d = 1.56 \times 1.25 \times 4.35 = 8.5$ d

Calculate the organic removal rate, q_{OBS}:

$$q_{OBS} = 1/(\theta_c^d \ Y_{NET}) \tag{6-5}$$

From Figure 2-10 at θ_c^d of 8.5 d, Y_{NET} is 0.24 g total VSS/g COD removed.

$q_{OBS} = 1/[(8.5)(0.24)] = 0.49$ g COD/g MLVSS/d

Determine reactor hydraulic retention time, t:

$$t = (S_0 - S_1)/(q_{OBS} \ X) \tag{6-6}$$

where:

> $S_0 = 187$ mg/L COD
> $S_1 = 20$ mg/L COD
> $X = 1,890$ mg/L MLVSS
> $t = (187 - 20)/(0.49 \times 1890) = 0.18$ d $= 4.3$ hr

Calculate reactor volume, V_N:

$$V_N = Q \ t$$

Since t in Equation 6-6 was computed based on the mg/L equivalent concentrations, the flow Q to compute the reactor volume is 18,925 m³/d (5 mgd) and not the actual flow of 21,056 m³/d (5.56 mgd) given in Table 2-16.

> $V_N = (18,925)(0.18) = 3,410$ m³ (0.90 Mgal)

8.5.3 Design Example No. 1 (continued)

2. *Determine various design parameters to check validity.*

Actual Retention time, t:

$t = (3,410/21,056) \times 24 = 3.9$ hr at Q of 21,056 m^3/d (5.56 mgd)

$t = 3.9/1.5 = 2.6$ hr at $Q_{max\ mo}$ of 31,584 m^3/d (8.34 mgd)

Food/Mass:

$F/M = (Q_{avg} \times BOD)/(V_N \times MLVSS)$

$F/M = (18,925 \times 97)/(3,410 \times 1,890) = 0.28$ g BOD/g MLVSS/d

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/d

At max. monthly:

$SNR_{min} = 0.069 \times PF$, where PF = 1.2 for influent TKN (Table 2-12)

$SNR_{min} = 0.069 \times 1.2 = 0.083$ g NH_4^+-N/g MLVSS/d

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:TKN ratios. In this case, $\theta_c = 8.5$ d, COD:TKN = 6.3, and CBOD:TKN = 3.3. If minimum rates calculated above are too low, increase the process design factor (DF) for sizing reactor (see Section 6.4.1.1).

3. *Sizing of anoxic zone for denitrification.* Size anoxic zone based on SDNR 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/d

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, or Eq. 4-22. In this example, determine the denitrification rate from Figure 8-30 (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/d at 20°C with wastewater as carbon source at anoxic F/M of 0.34 g BOD/g MLVSS/d at maximum monthly conditions.

Adjust nitrate removal rate using Eq. 4-25 at $\theta = 1.08$ (Table 4-2) for T = 15°C:

$SDNR_{15°C} = (0.09)(\theta^{T-20}) = (0.09)(1.08)^{15-20} = 0.061$ g NO_3^--N/g MLVSS/d

For maximum month where PF = 1.2: MLVSS required = 350 kg NO_3^--N/d x 1.2 PF ÷ 0.061 g NO_3^--N/g MLVSS/d from Table 2-12

= 6,885 kg (15,179 lb) MLVSS

Anoxic volume, V_{AN}:

$$V_{AN} = \frac{6,885 \text{ kg MLVSS} \times 10^3}{1,890 \text{ mg/L MLVSS}} = 3,643 \text{ m}^3 \text{ (0.96 Mgal)}$$
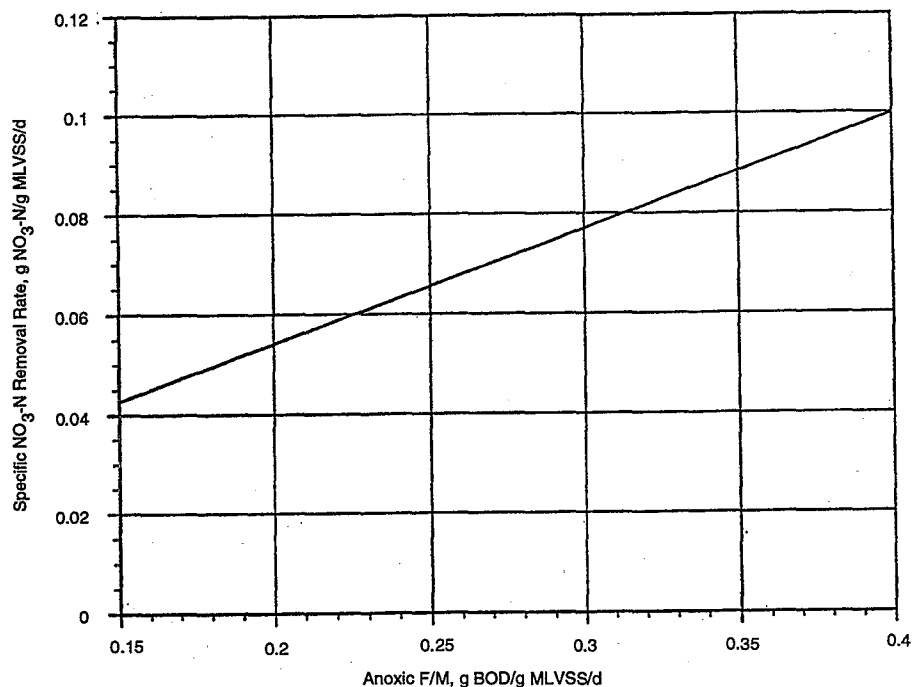
Figure 8-30. Denitrification rate as a function of anoxic F/M (adapted from 40).

Calculate hydraulic retention time in anoxic zone at actual flow:

At Q_{avg}, $t = (3,643/21,056) \times 24 = 4.2$ hr

At $Q_{max\ mo}$, $t = 4.2/1.5 = 2.8$ hr

$t = 4.2/1.5 = 2.8$ hr

System θ_c^d = aerobic θ_c^d + anoxic θ_c^d = 8.5 + [(3,643/3,410) × 8.5] = 17.6 d

4. *Determine RAS rate.*

To maintain MLSS = 3,000 mg/L at Q_{RAS} = 7,000 mg/L

$$Q_{RAS} = \frac{3,000 \times Q}{(7,000 - 3,000)} = 0.75Q$$

At Q = 21,056 m³/d (5.563 mgd), Q_{RAS} = 15,790 m³/d (4.17 mgd).

At $Q_{max\ mo}$ = 31,584 m³/d (8.34 mgd), Q_{RAS} = 23,690 m³/d (6.26 mgd).

5. Size anoxic internal recycle rate. From previous calculations of denitrification required and NO_3^--N returned to anoxic zone, the denitrification efficiency required equals 18.5/23.5 x 100 percent, which is 78.7 percent.

For single anoxic zone system:

$$\text{Denit. eff.} = \frac{Q_I + Q_{RAS}}{Q + Q_I + Q_{RAS}} \times 100\%$$

8.5.3 Design Example No. 1 (continued)

$$0.787 = \frac{Q_I + 0.75Q}{Q + Q_I + 0.75Q}$$

$$Q_I = \frac{[(0.787)(1.75) - 0.75]Q}{(1 - 0.787)}$$

$$Q_I = 2.95Q$$

At $Q_{max\ mo}$ = 31,584 m^3/d (8.34 mgd), Q_I = 93,160 m^3/d (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 = $7.14 \dfrac{mg\ CaCO_3}{mg\ NH_4^+\text{-}N}$ x 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 = $\dfrac{3.6\ mg\ CaCO_3}{mg\ NH_4^+\text{-}N\ reduced}$ x 18.5 = 67 mg/L as CaCO3

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/Mgal of anoxic volume, the minimum total hp required equals 48 where $V_{AN}^{'}$ = 3,643 m^3 (0.96 Mgal).

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 Section 6.4.2.2.*

$$\theta_c^d = I_A/S \tag{6-17}$$

where

I_A = VSS under aeration, and
S = VSS wasted daily

I_A = (1,890 mg/L MLVSS x 3,410 m^3)/1,000
  = 6,445 kg MLVSS (14,210 lb MLVSS)

Using (Eq. 6-17):

S = 6,445/8.5 = 758 kg (1,670 lb) VSS/d to be wasted

Solids contained in the effluent = 9 mg/L VSS from Table 2-15

Sludge wasting from RAS = 758 − (9) (18,925)/10^3 = 607 kg (1,340 lb) VSS/d

At MLVSS/MLSS = 63%:

Average WAS = 963 kg (2,120 lb) TSS/d

8.5.3 Design Example No. 1 (continued)

Determine WAS pumping rate at X_{RAS} = 7,000 mg/L:

$$\text{Average } Q_{WAS} = \frac{963 \ (10^3)}{7,000} = 138 \text{ m}^3/\text{d} \ (36,400 \text{ gpd})$$

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

$$\text{BOD}_5 \text{ Removed} = \frac{18,925 \text{ m}^3/\text{d} \times (97 - 3) \text{ mg/L}}{10^3} = 1,779 \text{ kg } (3,922 \text{ lb})/\text{d}$$

$$\text{NH}_4^+\text{-N oxidized} = \frac{18,925 \text{ m}^3/\text{d} \times 23.5 \text{ mg/L}}{10^3} = 445 \text{ kg } (981 \text{ lb})/\text{d}$$

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 (see Design Example 6-1 for more information).

Total oxygen demand = (1.1 x 1,779) + (4.6 x 445) = 4,004 kg (8,827 lb)/d

Peak Day Conditions (Peaking Factors From Table 2-12):

Total oxygen demand = (2.1 x 1.1 x 1,779) + (1.7 x 4.6 x 445)

$\qquad\qquad\qquad\qquad$ = 7,589 kg (16,730 lb)/d

Savings in O_2 demand with denitrification from wastewater BOD_5:

O_2 saved = 2.9 mg O_2/mg NO_3^--N x 18.5 mg/L NO_3^--N

$\qquad\qquad$ = 53.7 mg/L or 1,015 kg (2,237 lb)/d avg.

Oxygen demand with denitrification:

Avg O_2 demand = 4,004 − 1,015 = 2,989 kg (6,590 lb)/d

Peak day O_2 demand = 7,589 − (1.7)(1,015) = 5,864 kg (12,926 lb)/d

Mechanical aeration sizing

At 50 lb O_2/hp/d 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 Figure 8-31. 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. The following table summarizes reactor volumes at 10°C and 20°C for comparison. The volume changes reflect the impact of temperature in Equations 3-14 and 4-25, as well as the impact of θ_c on the observed yield, Y_{NET}, from Figure 2-10.

8.5.3 Design Example No. 1 (continued)

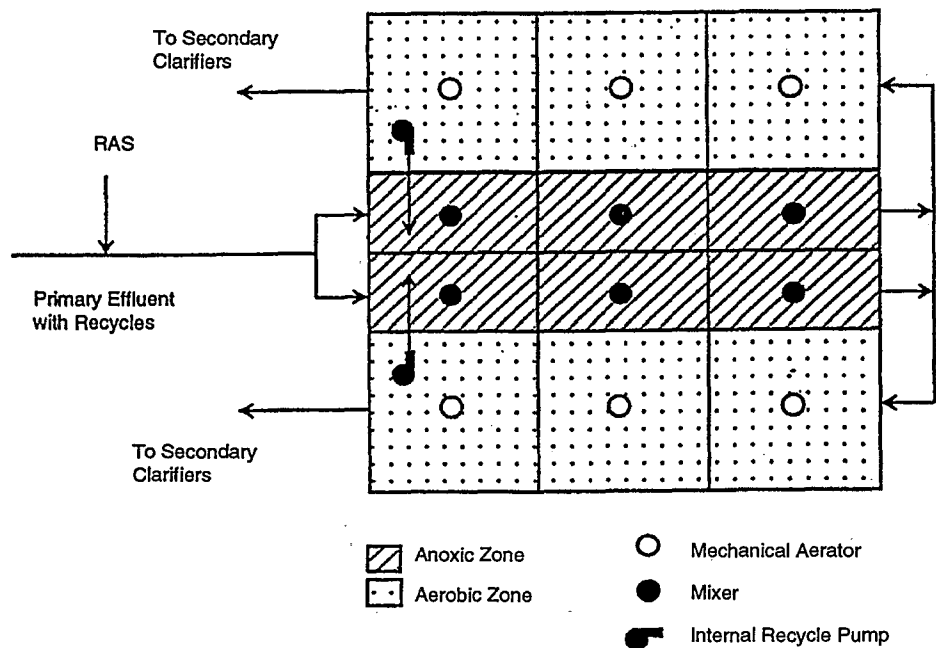| Volume, m^3	Design Temperature		
	10°C	15°C	20°C
V_{an}	5,370	3,640	2,270
V_{Nit}	4,980	3,410	2,380
V_{Tot}	10,350	7,050	4,650

Figure 8-31.   Design Example No. 1—Single anoxic zone system.

8.5.4 Design Example No. 2: Plant B with More Stringent Limits

The inlet wastewater characteristics (feed to secondary system) and secondary effluent characteristics from the mass balance in Table 2-16 are the same as in Example No. 1 for the purposes of this example. To meet the more stringent effluent limitations for SS, effluent filtration is required.

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 prior to 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 percent). 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 Figure 2-6 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 percent) will not be affected by the θ_c to remain consistent with the mass balance in Table 2-16.

Removal Requirements Across Secondary Process:

At average annual loadings (concentrations in mg/L equivalents):

BOD_5 Removed = 1,779 kg (3,922 lb)/d

$$\text{TN Removed} = \frac{18,925 \times (29.5 - 4.5)}{10^3} = 473 \text{ kg (1,043 lb)/d}$$

TN Removed in Waste Solids by Assimilation = 57 kg (125 lb)/d

TN Removed by Nitrification-Denitrification = 473 − 57 = 416 kg (917 lb)/d

1. *Sizing of nitrification zone (first aerobic zone.)* Following the same procedure used in Example No. 1, the reactor volume VN = 3,410 m³ (0.90 Mgal) using the same design factor. Since 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 (917 lb) NO_3^--N/d.

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.

8.5.4 Design Example No. 2 (continued)

Select Q_I = 450 percent or 4.5Q as practical limit. While Q_I can be higher, denitrification efficiency increases at a decreasing rate as shown in Figure 8-8.

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.

$$\text{Denitrification efficiency} = \frac{Q_I}{Q + Q_I + Q_{RAS}} \times 100\%$$

Q_{RAS} = 0.75 Q from Example No. 1

$$\text{Denitrification efficiency} = \frac{4.5Q}{Q + 4.5\,Q + 0.75\,Q} \times 100\% = 72\% \text{ maximum}$$

NO_3^--N Removed in first anoxic zone = 0.72 x 23.5 mg/L = 16.9 mg/L or 320 kg (706 lb)/d

NO_3^--N Removed required in second anoxic zone = 5.1 mg/L or 96 kg (213 lb)/d

Select SDNR as in Example No. 1. Since the amount of denitrification required in the first anoxic zone is similar to that in Example No. 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/d

$$\text{For maximum month, MLVSS required} = \frac{320 \text{ kg } NO_3^-\text{-N/d} \times 1.2 \text{ PF}}{0.061 \text{ g } NO_3^-\text{-N/g MLVSS/d}} = 6{,}295 \text{ kg } (13{,}880 \text{ lb})$$

$$V_{AN1} = \frac{6{,}295 \text{ kg MLVSS} \times 10^3}{1{,}890 \text{ mg/L MLVSS}} = 3{,}331 \text{ m}^3 \ (0.88 \text{ Mgal})$$

Calculate actual retention time

At Q of 21,056 m³/d (5.56 mgd), t = 3.8 hr

At $Q_{max\,mo}$ of 31,584 m³/d (8.34 mgd), t = 2.5 hr

3. *Determine RAS rate.*

Q_{RAS} = 0.75 Q similar to Example No. 1

4. *Internal recycle rate.*

As determined previously, Q_I = 4.5 Q

At maximum monthly flow, Q_I = 142,130 m³/d (37.5 mgd)

5. *Size second anoxic zone for denitrification with endogenous carbon.*

Determine $SDNR_2$ from Eq. 4-23 as a function of θ_c:

$SDNR_2 = 0.12 \times \theta_c^{-0.706}$ at 20°C

By trial and error, estimate that the system θ_c = 25 d.

$SDNR_2 = 0.12 \times 0.103 = 0.0124$ g NO_3^--N/g MLVSS/d

Adjust for T = 15°C using Eq. 4-25 and = 1.03 from Table 4.2:

$SDNR_2 = 0.0124 \times (1.03)^{-5} = 0.011$ g NO_3^--N/g MLVSS/d

NO_3^--N removal required in second zone = 96 kg (213 lb)/d, as previously determined.

8.5.4 Design Example No. 2 (continued)

$$\text{MLVSS required} = 96 \text{ kg/d} \frac{96 \text{ kg/d } NO_3^--N}{0.11 \text{ g } NO_3^--N/\text{g MLVSS/d}} = 8,727 \text{ kg (19,240 lb)}$$

$$V_{AN2} = \frac{8,727 \text{ kg MLVSS} \times 1.2 \times 10^3}{1,890 \text{ mg/L MLVSS}} = 5,541 \text{ m}^3 \text{ (1.47 Mgal)}$$

At Q of 21,056 m³/d (5.563 mgd), t = 6.3 hr

At $Q_{max \ mo}$, t = 4.2 hr

6. *Determine alkalinity requirements as in Example No. 1.*

7. *Determine mixing requirements as in Example No. 1.*

   50 hp/Mgal required for mixing in the anoxic zones

   $V_{AN1} + V_{AN2}$ = 3,331 + 5,541 = 8,872 m³ (2.35 Mgal)

   Total hp required = 118 hp

8. *Determine waste sludge as in Example No. 1.*

9. *Size post-aeration zone.*

Size for 0.5 hr detention time at $Q_{max \ mo}$ = 31,584 m³/d (8.34 mgd).

   V_{PA} = 660 m³ (0.17 Mgal)

10. *Calculate aeration requirements with fine pore diffused aeration system.*

Similar to Example No. 1 for carbonaceous oxidation and nitrification, but with post-aeration zone to raise DO from 0 to 2.0 mg/L.

   Total oxygen demand = 4,004 + 42 = 4,046 kg (8,920 lb)/d avg.

   $\qquad\qquad$ = 7,684 kg (16,940)/d peak day

Savings in O_2 demand with denitrification:

   O_2 savings = 2.9 mg O_2/mg NO_3^--N denit. x 22 mg/L NO_3^--N denit.

   $\qquad\qquad$ = 63.8 mg/L or 1,207 kg (2,660 lb)/d avg.

Oxygen demand with denitrification:

   Total oxygen demand = 4,046 − 1,207 = 2,839 kg (6,260 lb)/d avg.

   $\qquad\qquad\qquad$ = 7,684 − (1.7)(1,207) = 5,632 kg (12,416 lb)/d peak day

At 12.5 percent O_2 Transfer Efficiency (assumed at 4.6 m [15 ft] diffuser submergence):

   Peak air required = (5,632 kg O_2/d)/[(0.125)(0.28 kg O_2/m³ air)(1,440 min/d)]

   $\qquad\qquad\qquad$ = 112 m³/min (3,955 cfm)

Provide three blowers plus one standby

Blower Capacity = 38 m³/min (1,340 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 is similar to reactor configuration in Example No. 1, but add second anoxic zone and post aeration as shown in Figure 8-32. Nitrified flow is recycled from the end of the nitrification zone
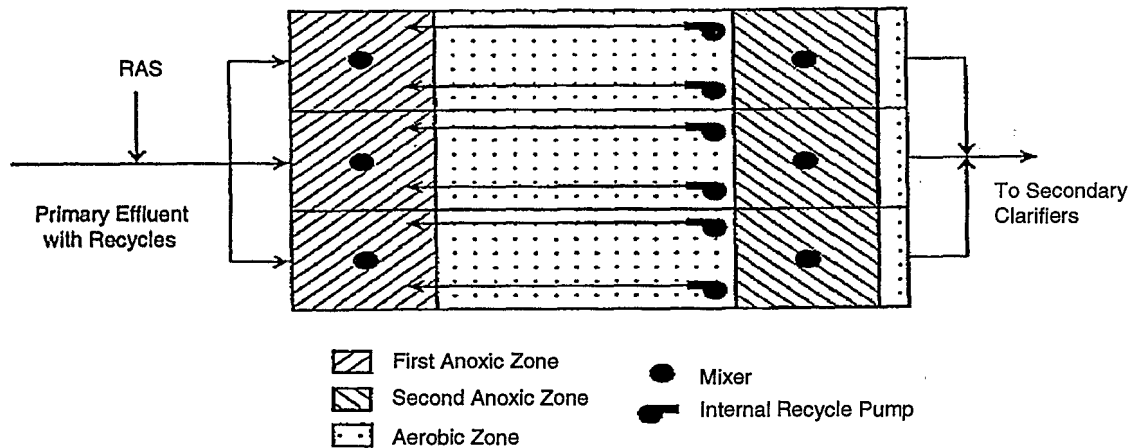
8.5.4 Design Example No. 2 (continued)

Figure 8-32. Design Example No. 2—Dual anoxic zone system.

to the head of the first anoxic compartment. The following table summarizes reactor volume requirements at 10°C, 15°C, and 20°C for comparison:

Volume, m^3	Design Temperature		
	10°C	15°C	20°C
V_{AN1}	4,950	3,330	2,110
V_{Nit}	4,980	3,410	2,380
V_{AN2}	6,450	5,540	4,670
V_{PA}	660	660	660
V_{Tot}	17,040	12,940	9,820

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 a TP limitation of 1.0 mg/L for the process configuration in this example, 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 1–2 hr 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 NO_3-N to denitrify) as this can cause an excess release of phosphates—a secondary release phenomenon, which occurs without storage of BOD.

8.5.4 Design Example No. 2 (continued)

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 post-aeration 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 $CaCO_3$.

8.5.5 Design Example No. 3: Plant A with Less Stringent Limits

From the mass balances in Table 2-15, the inlet wastewater characteristics for the less complex plant (raw influent plus recycles) and secondary effluent characteristics without nitrification-denitrification are as follows:

	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   Aerobic phase DO = 2.0 mg/L

Reactor MLSS = 3,000 mg/L   Secondary effluent NH_4^+-N = 1.0 mg/L

MLVSS = 58%   Secondary effluent Org-N = 2.0 mg/L

Reactor pH range = 7.0–7.6   Secondary effluent NO_3^--N = 5.0 mg/L

To meet seasonal limits of 2 mg/L NH_4^+-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.

8.5.5 Design Example No. 3 (continued)

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-percent 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 is as shown in Figure 2-5 for the less complex plant using cyclical 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 Rem. = (18,925 × 149)/1,000 = 2,820 kg (6,217 lb)/d

TN Rem. = [18,925 × (30.5 − 8.0)]/1,000 = 426 kg (939)/d

TN Rem. in Waste Solids by Assimilation = $\dfrac{18,925 \times 5.3}{10^3}$ = 100 kg (221 lb)/d

TN Rem. Required by Nitrification-Denitrification = 426 − 100 = 326 kg (719 lb)/d or 17.2 mg/L

NO_3^--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 No. 1, θ_c^d = 8.5 d for nitrification. Without separate digestion facilities, however, assume that sludge stabilization is required within the aeration tanks. For this example, to achieve stabilization, choose θ_c^d = 20 d for 10–20°C. Following the same procedure as in Example No. 1, the following calculation can be made:

q_{OBS} = 1/(20 × 0.27) = 0.185 g COD/g MLVSS/d

t = $(S_o-S_I)/(q_{OBS} \cdot X)$ = (290 − 20)/(0.185 × 3,000 × 0.58) = 0.84 d

V_{aer} = 0.84 × 18,925 = 15,900 m³ (4.20 Mgal); [(8.5/20) (15,900)] or 6,760 m³ (1.78 Mgal) required for nitrification

where Q = 18,925 m³/d was used to determine required aeration volume because t was computed based upon mg/L equivalent concentrations.

2. *Check design parameters.*

At Q = 19,379 m³/d (5.12 mgd), the actual retention time t = 19.7 hr

$F/M = \dfrac{18,925 \times 152}{15,900 \times 1,740}$ = 0.10 g BOD/g MLVSS/d

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 offcycle 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 percent of the anoxic phase and the wastewater carbon rate for 75 percent of the cycle.

Since the quantity of denitrification required is approximately the same as in the first anoxic zone of Example No. 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/d and $SDNR_1$ = 0.061 g NO_3^--N/g MLVSS/d. The weighed average is 0.049 g NO_3^--N/g MLVSS/d.

8.5.5 Design Example No. 3 (continued)

Denitrification required = 326 kg (719 lb)/d

For maximum month, MLVSS required = (326 x 1.2 PF)/0.049 = 7,984 kg (17,601 lb) MLVSS.

V_{AN} = (7,984 x 10^3)/1,740 = 4,589 m^3 (1.21 Mgal)

Check anoxic F/M:

F/M = (18,925)(152)/[(4,589)(1,740)] = 0.36, which is approximately equal to assumed 0.34 from Example No. 1 for trial-and-error solution.

Calculate actual hydraulic retention time for anoxic phase:

At Q = 19,379 m^3/d (5.12 mgd), t = 5.7 hr

At $Q_{max\ mo}$, t = 3.8 hr

4. *Determine ratio of anoxic and nitrification periods (cycling of on/off periods).*

V_{Nit} = 6,760 m^3 (1.78 Mgal)

V_{AN} = 4,589 m^3 (1.21 Mgal)

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 hr. 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 hr

On-cycle = 3.6 hr

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 No. 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 percent to account for the off-cycle. The aeration required for the final basins would be based on sludge stabilization requirements.

9. A typical flow configuration is shown in Figure 8-33. 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. The following summarizes reactor volumes at 10°C, 15°C, and 20°C for comparisons.

8.5.5 Design Example No. 3 (continued)

Volume, m³	Design Temperature		
	10°C	15°C	20°C
V_{Nit}	9,870	6,760	4,250
V_{AN}	6,400	4,590	3,100
V_{PA}	6,020	9,140	11,640
V_{Tot}	22,290	20,490	18,990

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.
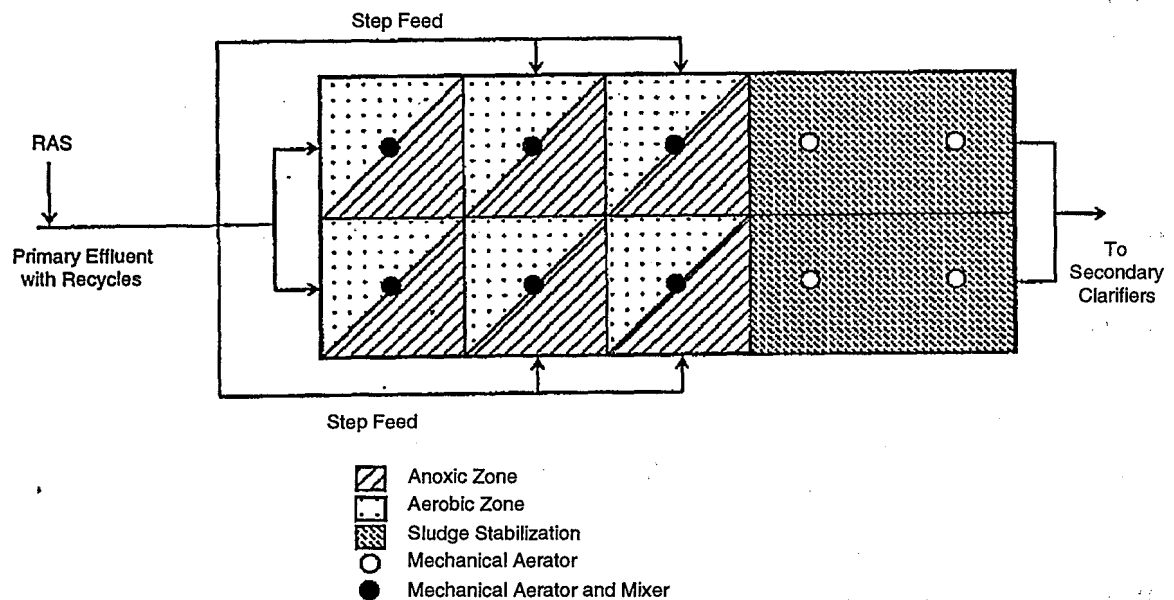

Figure 8-33. Design Example No. 3—CNR process.

308

8.5.6 Design Example No.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 No. 2 as shown in Figure 8-34. The procedures for sizing would be identical to those used in Example No. 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 as follows:

Volume, m^3	Design Temperature		
	10°C	15°C	20°C
V_{AN1}	5,040	3,390	2,150
V_{Aer}	9,870	6,760	4,250
V_{AN2}	8,370	7,190	6,060
V_{PA}	6,020	9,140	11,640
	29,300	26,480	24,100

RAS

Primary Effluent with Recycles

To Secondary Clarifiers

First Anoxic Zone
Second Anoxic Zone
Nitrification Zone
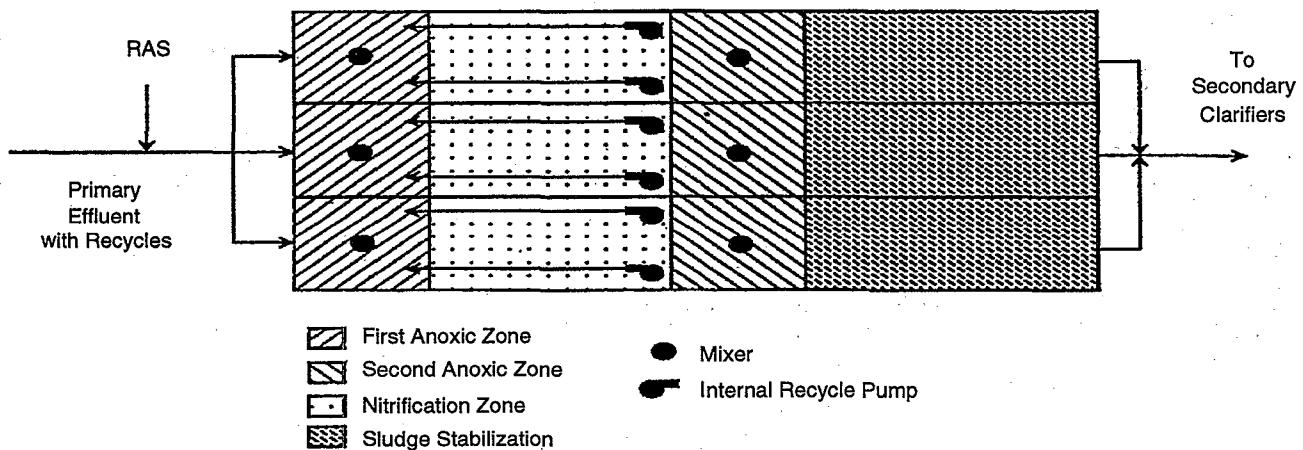Sludge Stabilization

Mixer
Internal Recycle Pump

Figure 8-34.   Design Example No. 4—Dual anoxic zone system.

8.6 References

When an NTIS number is cited in a reference, that document is available from:

National Technical Information Service
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Springfield, VA 22161
703-487-4650

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