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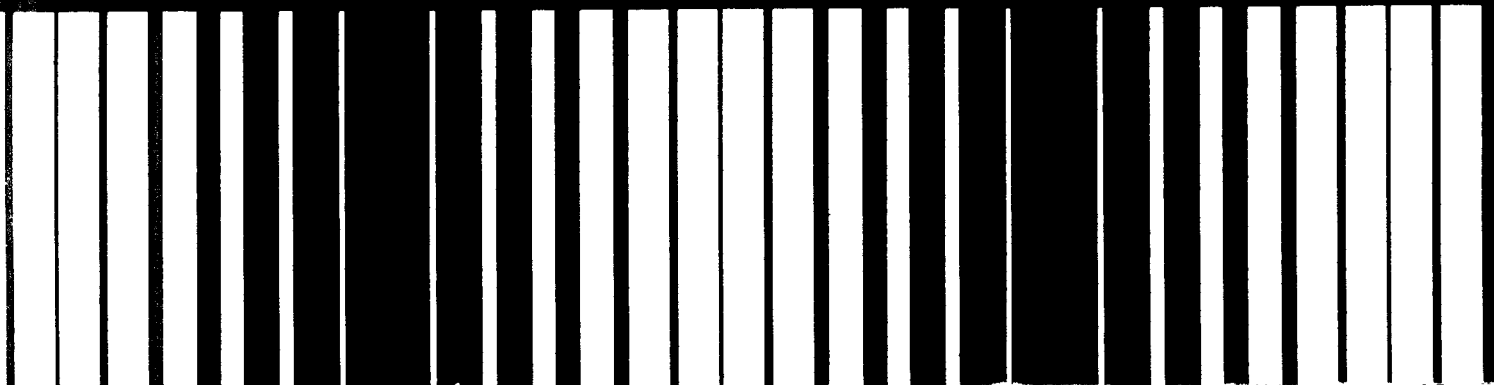
Technology Transfer

EPA 625 1-87 001



Design Manual

Phosphorus Removal



Design Manual

Phosphorus Removal

U.S. Environmental Protection Agency
Office of Research and Development

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Chapter 1

Introduction

1.1 Purpose

The Environmental Protection Agency has sponsored many research and demonstration studies at several cities in the past few years to advance the knowledge of phosphorus removal. Local and state governments and private industries have also contributed to this work. This manual is intended to summarize process design information for the best developed phosphorus-removal methods that have resulted from this governmental and private effort.

1.2 Scope

The sources and quantities of phosphorus in domestic wastewaters vary significantly. Industrial contribution, non-point source runoff, the use or nonuse of phosphate-bearing detergents, and other factors make generalization of expected wastewater phosphorus concentration impossible. The manual user should ascertain the phosphorus concentrations (actual or expected) for the specific wastewater in question.

This manual discusses several proven phosphorus-removal methods, including phosphorus removal obtainable through biological activity as well as chemical precipitation techniques.

Biological phosphorus removal was not included in the previous version of this manual and represents a major addition.

Appropriate chemistry for phosphorus removal by chemical addition is presented where appropriate for illustrative purposes.

The use of lime as a chemical precipitant for phosphorus removal, which received major treatment in the previous version, is not covered in this manual due to its loss of popularity as a phosphorus-removal technique. The reasons for current infrequent use of lime are discussed in Chapter 4. The user can refer to several other documents for detailed information and design criteria for lime addition, including the previous version of this manual (1).

Treatment methods in which phosphorus removal occurs, but is not a principal objective, are also omitted. The latter group of processes includes such technologies as ion exchange, reverse osmosis and other demineralization treatments which at present are more closely associated with wastewater renovation and reuse than with pollution control. These will be included in updated versions of the manual when appropriate.

The information included was obtained from the available literature, progress reports from demonstration studies, and private communications with investigators actively working in the field. Design guidelines have been developed from these sources.

The information contained in this manual is oriented toward design methods and operating procedures for removal of phosphorus from wastewater.

Cost information from actual phosphorus-removing installations is presented when available. Planning level cost estimates are also included.

1.3 Using the Manual

Chapter 2 presents a recommended approach to selecting a phosphorus removal strategy. This approach identifies the required effluent phosphorus limits and screens potential phosphorus-removal techniques to identify those processes capable of meeting the specified requirements.

The screening process is a step-by-step procedure that identifies the information required to make the engineering judgements necessary at each step of the process. After applying the screening methodology, the manual user can go to the specific chapter in the manual dealing with the potential phosphorus-removal processes available for his specific case:

Chapter 3 - biological phosphorus removal.

Chapter 4 - phosphorus removal by mineral addition.

Chapter 5 discusses all aspects of sludge handling associated with sludges generated from phosphorus-removing facilities.

1.4 Reference

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

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

1. *Process Design Manual for Phosphorus Removal*. EPA 625/1-76-001a, NTIS No. PB-259150, U.S. Environmental Protection Agency, Center for Environmental Research Information. Cincinnati, OH, 1976.

Chapter 2

Selecting a Phosphorus Removal Strategy

2.1 Description of Approach

The approach described herein for selecting a phosphorus removal system identifies the required effluent phosphorus limits, and then screens phosphorus removal processes to identify those processes capable of meeting specified requirements. All alternative phosphorus removal technologies are considered initially, but non-applicable technologies are eliminated through a sequential selective screening process.

The selective screening methodology is a step-by-step procedure that identifies information required to make the necessary engineering judgments at each step of the process. Brief descriptions of the various chemical and biological phosphorus removal processes and the advantages and disadvantages of each process have been provided. The effectiveness of the selective screening process is dependent upon the amount and level of detail of the initial information used.

For each specific phosphorus removal alternative considered, the sections of the manual dealing with that process should be read carefully before final selection of a phosphorus removal system. Other sources of background information include manufacturers or proprietors of phosphorus removal technologies, and site visits to operating phosphorus removal plants. The process should not be considered completely rigorous, since some subjective judgments are required. It is intended as a comprehensive overall guide to the selection of an appropriate phosphorus removal process.

The selection procedure must consider all aspects of the phosphorus removal process including its impact on plant performance, operations and maintenance. Important factors are: a) degree of phosphorus removal required, b) size of plant, c) impact on sludge handling, d) permanent or temporary nature of phosphorus removal requirement, e) total cost and f) impact on operation and maintenance.

The system screening process is presented in Section 2.4. The basic information needed to apply the screening procedure is discussed in Section 2.2,

and a brief description of the state-of-the-art phosphorus removal technologies is presented in Section 2.3.

2.2 Information and Monitoring Data Required

The information and monitoring data required for selecting a phosphorus removal system are described below. Information required for new plants is somewhat different than for existing facilities. Most phosphorus removal systems will be retrofitted to existing systems due to imposition of new permit requirements. In either case, knowledge of what information is required to make a cost effective selection is critical in being able to utilize the screening methodology presented later.

2.2.1 New System Data Requirements

2.2.1.1 Effluent Discharge Requirements

The first step taken in evaluating alternative phosphorus removal processes for a new facility is to determine the phosphorus discharge limitations. These limits should be determined in as detailed a manner as possible to define the daily, weekly, monthly and possibly seasonal phosphorus limits. Determination of whether an existing facility will be phased out due to regionalization or plant consolidation should also be ascertained to ensure that the time frame used in the cost analysis reflects the expected plant lifetime.

Phosphorus limits may be set as a minimum percent removal, as a specific effluent concentration or on a mass per day basis. In addition to the phosphorus discharge limitations, the permit limits for BOD₅, TSS, pH, NH₄-N or total N must be known. The necessity to remove ammonia and/or total-nitrogen from the effluent can have a significant impact on the selection of the phosphorus removal process. The information normally required is shown in Table 2-1.

2.2.1.2 Wastewater Characteristics

Once effluent limitations have been determined, the next step is to develop the information on wastewater characteristics which may affect the choice of phosphorus removal alternatives. For a new facility,

Table 2-1. Information Required for Discharge Limits

Parameter	Weekly	Monthly	Seasonal
Total P, mg/l			
Total P, kg/d			
Total P, % removal			
BOD ₅ , mg/l			
TSS, mg/l			
NH ₄ -N, mg/l			
Total N, mg/l			
pH, units			

Table 2-2. Information Required for Influent Wastewater Characteristics

Parameter	Daily ¹	Weekly	Monthly
Flow			
Average, m ³ /d			
Maximum, m ³ /d			
Peak/Average Ratio			
Total P, average, mg/l			
Soluble P, average, mg/l			
BOD ₅ , average, mg/l			
SBOD, average, mg/l			
TSS, average, mg/l			
NH ₄ -N, average, mg/l ²			
Total N, average, mg/l ²			
pH, units			
Temperature			
Alkalinity, average, mg/l			
TBOD:Total P ratio			

¹ Seasonal column should be added for biological phosphorus removal facilities.

² Where effluent discharge permit limits Total N or NH₄-N.

the most important wastewater characteristics are listed in Table 2-2. In addition, any significant contribution from individual sources should be characterized. The effect of these variables on alternative phosphorus removal processes is summarized in Section 2.4. Detailed explanations are provided in Chapters 3 and 4 for the specific processes.

The best data on anticipated plant flows and wastewater characteristics are from plant influent monitoring data. If data on nutrients, alkalinity, etc. are not available, it is recommended that data from the literature be used for facilities of similar size and service characteristics. Actual characterization data from the wastewaters to be treated are much preferred, and every effort should be expended to

collect representative information. The importance of adequate wastewater characterization cannot be over-emphasized. The cost of sampling and analysis required for proper wastewater characterization is almost always offset by more efficient system design and lower total costs for phosphorus removal.

Alkalinity and pH data are site specific and must be determined for each facility. Where plant influent data are not available, alkalinity can be best obtained from the water supply sources serving the contributing wastewater generators.

2.2.1.3 Other Information Required

Other information that may have a significant influence on the phosphorus removal alternatives are:

1. Sludge disposal alternatives.
2. Service area characteristics including industrial contributions.
3. Plant size and location; available land.
4. Facility design lifetime.
5. Local availability and cost of chemicals.

Smaller plants in more rural settings generally offer a wider range of alternatives because of the reduced constraints of sludge handling. Chemical alternatives generally require less space but exhibit higher operation and maintenance costs. Biological alternatives require more careful analysis of total treatment requirements including the proper selection of a basic BOD removal process.

Sludge generation can be a significant factor influencing the phosphorus removal alternative selected. Many communities are limited in their choice of sludge disposal options due to local or state regulations, air quality requirements or biases on the part of local decision makers and/or local concerned citizens. Because of the importance of sludge handling cost as a part of the total treatment costs, new phosphorus removal projects must consider sludge handling as an integral part of the total system analysis.

The typical planning period for new wastewater treatment facilities is 20 years. However, phosphorus removal standards may be established for different time frames. Chemical alternatives may be more cost effective for shorter term design periods, while biological alternatives with higher capital costs may be more appropriate for longer term projects.

Wastewater characteristics are determined to a large degree by the composition of the contributing sewer dischargers. The breakdown of the system wastewater contributors into residential, commercial and industrial discharges can be made from community records and/or planning studies. Where significant industrial contributors are identified, the impact of their wastes on the wastewater composition

should be reviewed for its effect on the various phosphorus removal alternatives selected for consideration.

2.2.2 Existing System Data Requirements

Incorporation of phosphorus removal technology into existing facilities presents a somewhat more complicated initial analysis than that required for new systems. Existing systems may offer opportunities for cost effective incorporation of some of the available phosphorus removal technologies, but at the same time may present severe constraints to the use of others. The more important considerations for existing facilities are:

1. Whether or not the existing facility is meeting or can be easily upgraded to meet the non-phosphorus effluent limits.
2. Whether or not the existing hydraulic capacity is adequate for the proposed design flows.
3. Whether or not the age and condition of the existing plant justifies the inclusion of a new phosphorus removal technology.
4. Whether or not the existing facility must be upgraded (to improve effluent quality) or expanded, or both.
5. Whether or not the basic form of biological treatment, i.e., fixed film or suspended growth, is compatible with phosphorus removal technologies capable of meeting the required effluent phosphorus limitations.

Once these basic questions have been addressed, the analysis may proceed in the same manner described earlier for new facilities.

1. Type, capacity and efficiency of existing unit operations in service:
 - a. Liquid stream
 - hydraulic capacity
 - biological capacity
 - aeration capacity
 - modifications planned or underway
 - b. Solids stream
 - thickening and dewatering capacity
 - stabilization capacity and efficiency
 - ultimate disposal capacity
 - compatibility of sludge handling with applicable phosphorus removal technologies

2. Unit operation performance for specific parameters, including:

Total P
TSS

BOD₅
Total N
NH₄-N

3. Unit operations known to be obsolete or to require replacement.
4. Unused or abandoned plant capacity that could be brought into service.
5. Discharge points of recycle streams from existing sludge handling systems; and loadings associated with these streams for the above parameters.
6. Sludge disposal options available with existing sludge handling system; loading capacities and efficiency of all sludge handling processes. Capacity analysis of solids handling systems should note the percent of time units are in service and the percent of existing operating capacity that is being used.

Personnel requirements for operation of existing equipment should be developed so that effects of adding new operations or expanding operating hours of existing equipment can be used in evaluating the additional costs of various phosphorus removal alternatives. The level of sophistication of operations for the applicable phosphorus removal technologies should be compared to the existing plant operation requirements. For example, if existing plant performance is inadequate due to deficient operations, the incorporation of a biological phosphorus removal technology requires substantial upgrading of operator staffing and competence level.

2.3 Possible Phosphorus Removal Alternatives

Detailed descriptions of phosphorus removal technologies and the attendant sludge handling alternatives are presented in Chapters 3 through 5. This section briefly summarizes the various phosphorus removal technologies. Simplified process flow schematics, along with expected performance ranges and the major advantages and disadvantages of each process, are discussed below.

2.3.1 Chemical Addition Alternatives

2.3.1.1 Metal Salt Addition (Aluminum and Iron)

Metal salts of aluminum and iron added to wastewater react with phosphates to form insoluble aluminum or iron phosphate precipitates. These compounds include aluminum sulfate (alum), sodium aluminate, ferric chloride, ferrous chloride, and ferrous sulfate. They are generally added upstream of either the primary clarifier or the secondary clarifier. In some cases metal salts are added to both primary and secondary processes. Figure 4-1 shows alternative

metal salt addition flow schematics. The chemicals may also be added separately to a tertiary clarifier. The quantity of metal salt added is determined by the concentration of phosphorus species in the influent wastewater and effluent discharge permit requirements.

Systems with metal salt addition can achieve 80-95 percent total phosphorus removal. For effluent limitations of 1.0 mg/l total phosphorus, metal salt addition with conventionally designed clarifiers is acceptable. Effluent limitations of 1.0 mg/l total phosphorus can be met with metal salt addition and efficient clarification to assure effluent TSS of less than 15 mg/l. To consistently meet total phosphorus discharge limitations of 0.5 mg/l, filtration of secondary effluent will most likely be necessary.

Determination of the best point or points of chemical addition is best determined by full-scale plant testing. Jar tests using the different metal salts can provide sufficient information needed to conduct cost estimating and evaluation of the relative impact of sludge production on sludge handling and disposal processes. Pilot or full-scale testing is recommended to develop detailed design criteria where stringent discharge requirements (e.g., 0.5 mg/l total phosphorus) are imposed.

Advantages and disadvantages of metal salt addition for phosphorus removal are summarized in Table 2-3.

2.3.1.2 Lime Addition

Lime is used to remove phosphorus by addition either to the primary clarifier or to the effluent from the secondary clarifier in a separate tertiary unit.

Phosphorus removal with lime is basically a water softening process and the quantity of lime required to remove the phosphorus is dependent on the alkalinity of the wastewater rather than on the phosphorus content.

Lime removal systems are either low-lime (single stage) where pH is kept below 10.0 and which can achieve 1.0 mg/l effluent total phosphorus levels or a two-stage high-lime process that raises the pH to 11.0-11.5 and is used to achieve very low (<1.0 mg/l) effluent total phosphorus concentrations. High-lime treatment uses more lime and also requires recarbonation to reduce the wastewater pH before discharge to a downstream biological unit or discharge from the plant. The effluent phosphorus levels from a two-stage tertiary lime process with filtration can meet total phosphorus effluent limits as low as 0.1.

The lime addition process produces a substantial amount of additional sludge, even greater than with metal salt addition.

Lime addition systems are usually pH controlled and entail lime storage, feed, and mixing units. This equipment often requires considerable maintenance.

Lime can be reused by calcining the lime sludge. This step is only applicable at larger plants due to the high capital and operating costs for the recalcining process. Even with recalcining, about 20-30 percent of make-up lime is required.

Lime addition is seldom practiced in the United States today due to the high chemical usage, problems associated with handling lime, and the large volume of sludge generated from lime addition systems. Advantages and disadvantages of lime addition for phosphorus removal are summarized in Table 2-4.

2.3.2 Biological Phosphorus Removal Alternatives

Biological phosphorus removal is a recently developed technique of designing suspended growth activated sludge systems to remove soluble phosphorus from wastewater. Six variations on this phenomenon are described in Chapter 3 of this manual. These alternatives are:

- Phostrip process
- Modified Bardenpho process
- A/O process
- UCT (University of Capetown) process
- Sequencing Batch Reactor (SBR) process
- Operationally modified activated sludge processes

Figures 3-4, 3-5, 3-6 and 3-7 are flow diagrams of these processes. Detailed descriptions of the various biological processes and the theories and mechanisms of operation can be found in Chapter 3. Case histories of various facilities are also presented.

Phostrip is the only biological phosphorus removal process that incorporates an anaerobic zone in the sludge recycle system. The Phostrip process takes a sidestream from the return activated sludge and subjects it to anaerobic conditions in a separate anaerobic tank before returning the sludge to the aeration basin. Maintenance of the activated sludge in the anaerobic state leads to phosphorus release, and when the sludge is returned to the aeration basin or reaerated, the activated sludge biota take up an excess amount of phosphorus during the growth process. Lime is used to precipitate the phosphorus released in the anaerobic tank. Since only 20 to 30 percent of the plant flow passes through the anaerobic tank, the quantity of lime required is much less than in a mainstream lime addition system, and less sludge is produced.

The Modified Bardenpho process and the UCT process are designed to remove both nitrogen and phosphorus.

Table 2-3. Advantages and Disadvantages of Metal Salt Addition for Phosphorus Removal

Advantages	Disadvantages
1. Reliable, well documented phosphorus removal technique. Most popular process used in the United States.	1. Chemical costs higher than for biological phosphorus removal systems which require little or no metal salt addition
2. Chemical costs can be reduced substantially if waste pickle liquors (ferrous chloride or ferrous sulfate) are available and can be used.	2. Significantly more sludge produced than with wastewater treatment process without metal addition; may overload existing sludge handling equipment; higher sludge treatment and disposal costs.
3. Chemical usage requirement is basically dependent on total phosphorus concentration of wastewater and required effluent levels.	3. Sludge produced generally does not dewater as well or as easily as conventional wastewater treatment plant sludges where metal salts are not added.
4. Controls required for phosphorus removal are fairly simple and straightforward.	
5. Relatively easy and inexpensive to install at existing facilities	
6. Sludge produced can be processed in same manner as in non-phosphorus removal systems.	
7. Primary clarifier metal addition can reduce organic load to secondary unit by 25-35 percent.	
8. Effluent phosphorus levels can be controlled by metal salt dosages to maximum efficiency levels.	

Table 2-4. Advantages and Disadvantages of Lime Addition for Phosphorus Removal

Advantages	Disadvantages
1. Simple process control, as lime dosage paced by pH control. Lime dosage required does not vary with phosphorus concentrations, only alkalinity of wastewater.	1. High chemical costs for wastewater facility with hard (high alkalinity) waters.
2. Very high phosphorus removals achievable with high lime process.	2. More sludge produced than for any other phosphorus removal process.
3. Many heavy metals, such as chrome, nickel, etc., are effectively removed.	3. Equipment requirements and maintenance costs for lime storage, feeding, and handling equipment are extremely high.
4. Primary lime addition reduces organic load to biological treatment units.	4. High capital and operating costs; not widely used in the United States today.
	5. Additional recarbonation step required for high lime process.

The A/O process is primarily designed for phosphorus removal but it can also be designed to accomplish both phosphorus removal and nitrification.

Biological phosphorus removal has been achieved in existing plug flow activated sludge plants by modifying the aeration practice at the head end of the aeration tanks. This modification consists of shutting off the air or aerators at the head of the tank to promote an anoxic or anaerobic zone before the wastewater and return sludge are aerated.

Phosphorus removal in an SBR has been accomplished in limited full-scale tests by modifying the sequencing periods and aeration schedules to provide the needed anaerobic/aerobic mixing periods.

The sludge quantities produced by all the biological phosphorus removal processes, with the exception of

Phostrip, are not any greater than those produced from conventional suspended growth systems. Processing of the biological phosphorus sludges, however, requires care that the sludge processing steps do not result in phosphorus resolubilization, and return of soluble phosphorus to the system.

The degree of phosphorus removal from biological phosphorus removal processes (except Phostrip) is dependent on the influent BOD:P ratio of the wastewater. Within the proper BOD:P ranges, the Modified Bardenpho, A/O, UCT and operationally modified activated sludge processes can achieve 1.0-2.0 mg/l total phosphorus (TP) level in the final effluent. A TBOD:TP ratio of at least 20:1 (SBOD:soluble P of 12:1 to 15:1) is usually required to meet these limitations. Achievement of lower effluent total phosphorus concentrations normally requires efficient clarification to achieve less than 20

Table 2-5. Advantages and Disadvantages of Biological Phosphorus Removal Processes

Advantages	Disadvantages
1. Sludge quantities generated by biological phosphorus removal processes are comparable to sludge production from conventional activated sludge systems.	1. In all but Phostrip, phosphorus removal performance is controlled by the BOD:TP ratio of the wastewater.
2. Can be implemented directly at existing plug flow activated sludge plants with little or no equipment changes or additions, provided that the plant has sufficient capacity.	2. Requires highly efficient secondary clarifier performance to achieve 1 mg/l total phosphorus.
3. Can utilize existing sludge handling equipment for plants retrofitted with biological phosphorus removal process if phosphorus is not solubilized and returned to the plant during sludge handling.	3. Not easily retrofitted into fixed film biological systems.
4. Little or no chemicals or chemical handling equipment required except for Phostrip process or for effluent polishing.	4. Potential for phosphorus release in sludge handling system. Recycle streams must be low in phosphorus content.
5. Phosphorus removal can be accomplished together with ammonia nitrogen or total nitrogen removal at virtually no additional operating cost with some of the processes.	5. Standby chemical feed equipment may be necessary in case of loss of biological phosphorus removal efficiency.
6. For some of the processes, better control of filamentous organisms in the activated sludge system is possible.	

mg/l TSS concentrations or the use of tertiary filters. Discussion of the effect of BOD:P ratios and other factors on phosphorus removal is contained in Section 3.3.5.

In all these systems, pilot testing is highly recommended to determine what performance levels can be achieved for the specific wastewater to be treated. Advantages and disadvantages of biological phosphorus removal systems are summarized in Table 2-5.

2.4 Phosphorus Removal System Selection Strategy

The strategy that is described below for selecting a phosphorus removal system is a selective screening process whereby all phosphorus removal alternatives are initially considered, and then are screened against various sets of criteria that impact the choice of a system. At each step in the four step process, alternatives are evaluated against a corresponding application matrix (or set of matrices) which shows the applicability of the process for various sets of conditions. Non-applicable technologies are rejected at each step with the selected alternative being chosen as a result of the fourth step of the process, which is the final cost-effectiveness analysis. A flow diagram of the overall selection strategy is depicted in Figure 2-1.

It should be emphasized that this approach is intended as a guide for assisting the engineer in the overall selection process. The approach is not rigorous, since subjective information is used in the selection/rejection process and procedures for reiterating through the process are not provided.

The four steps of the selective screening process are described below.

Step 1:

Categorize the facility as to whether it is an existing or a new plant, and whether the effluent nutrient discharge limitations are for phosphorus only, or for phosphorus plus nitrogen. In general, for new facilities, all possible alternatives should initially be considered, although some may have greater apparent applicability than others. For existing facilities, some alternatives can justifiably be removed from further consideration.

For example, for a trickling filter plant with sufficient design capacity that must meet a new phosphorus limitation, it is unlikely that a new biological phosphorus removal system would be competitive with metal salt addition to the primary clarifier. However, if the same plant were required to meet both phosphorus and nitrogen limitations, a new biological suspended growth system designed for N and P removal may be more cost effective than modifying the existing plant to achieve nutrient removal. The applicability criteria matrix for Step 1 screening is shown as Table 2-6.

Step 2:

Apply P-removal process capabilities and determine which processes can meet phosphorus limitations.

For those plants with phosphorus and nitrogen limitations, the ability of the process to meet nitrogen limitations will also be considered. In conducting a study of plants having both phosphorus and nitrogen effluent limitations, an additional screening step should be made to include other nitrogen removal processes that can be used independent of the

Figure 2-1. Phosphorus Removal System Selection Strategy.

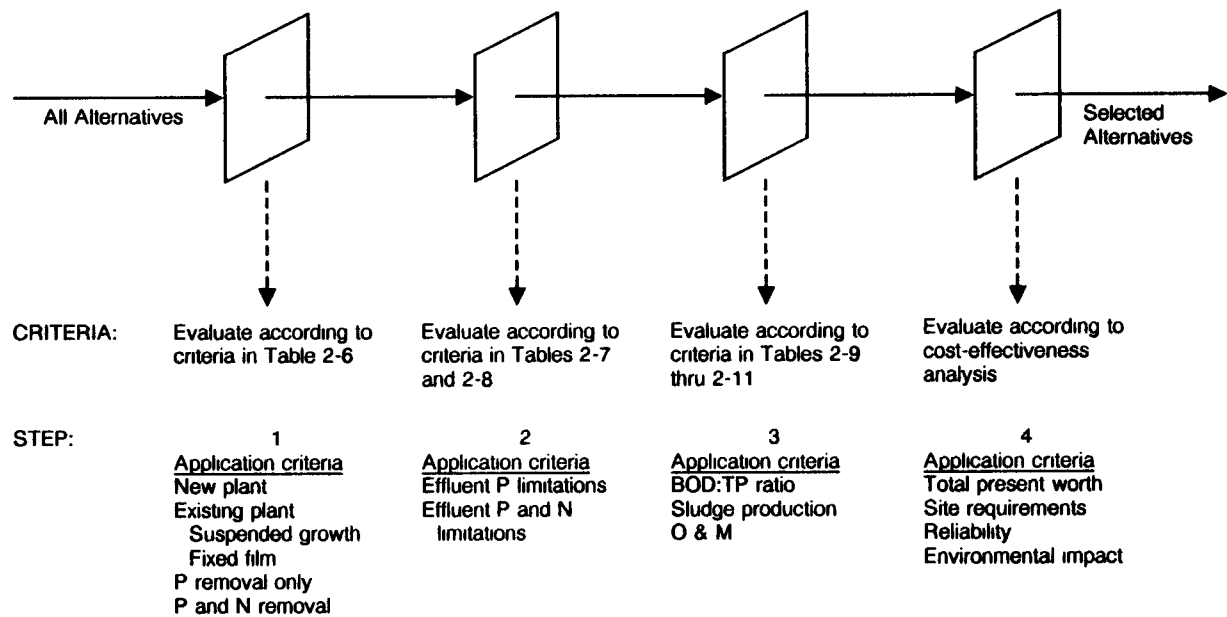


Table 2-6. Application Criteria Matrix - New or Existing Facility; Type of Nutrient Removed

Process	P Removal Only			P plus NH ₄ or N Removal		
	New Facility	Existing Susp. Growth	Existing Fixed Growth	New Facility	Existing Susp. Growth	Existing Fixed Growth
(A) Phostrip	Y	Y	N	Y	M	Y*
(B) Mod. Bardenpho	E	E	N	Y	Y	N
(C) A/O	Y	Y	N	Y	M	Y*
(D) SBR	Y	Y	N	Y	Y	N
(E) UCT	E	E	N	Y	Y	N
(F) Mod. A.S.	Y	Y	N	Y	M	Y*
(G) Metal Salt	Y	Y	Y	Y	M	M
(H) 1-Stage Lime	Y	Y	Y	Y	M	Y*
(I) 2-Stage Lime	Y	Y	Y	Y	M	Y*

Y - Applicable
N - Not Applicable

E - May be Applicable, but exceeds treatment requirements.
M - May be Applicable, but N or NH₄ removal step required.
Y* - Applicable where existing fixed film unit can be used for nitrification.

phosphorus removal process, i.e. consider alum for phosphorus removal and breakpoint chlorination, or a separate nitrification tower to meet the P and N limitations.

In addition, effluent limitations such as BOD₅ and suspended solids as well as other site specific requirements not specifically listed as "application criteria" may lead to rejection or inclusion of some P-removal processes.

To help select the processes which can meet the various phosphorus, or phosphorus and nitrogen limitations, Tables 2-7 and 2-8 are used. These tables are matrices showing the ability of the systems to meet various phosphorus limits as well as being able to meet effluent nitrogen limits. These matrices are intended to be used as guidelines and are not intended for rigid process selection. Specific data on the nitrogen removal/conversion characteristics of the biological phosphorus removal processes must be studied carefully before determining whether a specific process should be included or eliminated as a possible alternative.

Step 3:

Step 3 consists of screening the alternative processes that have been found to be capable of meeting the effluent phosphorus or phosphorus and nitrogen limitations with the applicability criteria in Tables 2-9 through 2-11. The objectives of this screening are to eliminate those processes which are not suitable for further consideration. Where a process is shown to be marginal and other processes are shown to be capable of meeting the effluent limitations, the marginal processes should also be carried through to step 4.

The Application Criteria Matrices shown as Tables 2-9 through 2-11 have been developed to provide a means of quickly determining the factors which are pertinent for screening the different phosphorus removal processes. As in the previous tables, these matrices are intended to provide a means of evaluating the impact of various factors on the different phosphorus removal processes. In using the matrices the user should read the applicable sections of the manual which describe the processes in detail. A brief description of the use of the matrices follows.

Table 2-9, Effect of Influent TBOD:TP Ratio on Process Applicability, is based on research that indicates plants with a TBOD:TP ratio less than 20 may have difficulty achieving final effluent total phosphorus concentrations of 1.0-2.0 mg/l. Specific sampling data on the variation of TBOD:TP ratio and/or pilot-plant studies for a specific facility may be necessary to define the capabilities of the biological phosphorus removal process under consideration.

Table 2-10, Sludge Production, indicates the effect on sludge production of the different phosphorus removal systems for typical process applications. Operation and maintenance requirements for increased sludge production are included in the ratings for the processes. More specific data on sludge production for the specific processes are presented in Chapter 5.

Table 2-11, O & M Effects, shows the difference in O & M requirements for the different processes. The O & M comparison does not include sludge handling and disposal since this aspect is covered in Table 2-10.

Step 4:

Step four consists of developing the capital, operation and maintenance, and total present worth costs for all applicable alternatives. Non-monetary factors are also considered at this point, including:

1. Site requirements.
2. Reliability.
3. Environmental Impacts.
4. Operator skill level required for successful operation.

The results of the comprehensive cost effectiveness analysis, which evaluates costs of screened alternatives against appropriate non-monetary factors, results in the selection of a system which meets project objectives at the lowest present worth cost.

Table 2-7. Application Criteria Matrix - Ability of Process to Meet Effluent Phosphorus Limitations

Process	0.5 mg/l Effluent TP				1.0 mg/l Effluent TP				2.0 mg/l Effluent TP			
	Alone	w/M.S.	w/F	w/M.S. & F	Alone	w/M.S.	w/F	w/M.S. & F	Alone	w/M.S.	w/F	w/M.S. & F
(A) Phostrip	N	N	N	Y	Y	Y	Y	Y	Y	Y	Y	Y
(B) Bardenpho	N	N	N	Y	M	Y*	M	Y	Y	Y	Y	Y
(C) A/O	N	N	N	Y	M	Y*	M	Y	Y	Y	Y	Y
(D) SBR	N	N	N	Y	M	Y*	M	Y	Y	Y	Y	Y
(E) UCT	N	N	N	Y	M	Y*	M	Y	Y	Y	Y	Y
(F) Mod. A.S.	N	N	N	Y	M	Y*	M	Y	Y	Y	Y	Y
(G) Metal Salt	N	-	Y	-	Y*	-	Y	-	Y	-	Y	-
(H) 1-Stage Lime	N	N	Y	Y	Y*	M	Y	Y	Y	Y	Y	Y
(I) 2-Stage Lime	N	M	Y	Y	Y*	M	Y	Y	Y	-	Y	-

Ability of alternatives (B)-(F) to meet effluent limits alone based on TBOD:TP ratio being above 20.

N - Can not meet effluent limits.

M - Marginal for meeting effluent limits.

Y - Will meet effluent limits.

Y* - Will meet effluent limits with highly efficient clarification.

M.S. - Metal salt addition to secondary clarifier.

F - Filtration of secondary clarifier effluent.

M.S. & F - Metal salt addition to secondary clarifier and secondary clarifier effluent filtration

Table 2-8. Application Criteria Matrix - Ability of Process to Remove NH₄ or TN

Process	0.5 mg/l Effluent TP		1.0 mg/l Effluent TP		2.0 mg/l Effluent TP	
	NH ₄	TN	NH ₄	TN	NH ₄	TN
(A) Phostrip	N/A	N/A	Y	N	Y	N
(B) Bardenpho	N/A	N/A	Y	Y	Y	Y
(C) A/O	N/A	N/A	Y	N	Y	-
(D) SBR	N/A	N/A	Y	Y	Y	Y
(E) UCT	N/A	N/A	Y	Y	Y	Y
(F) Mod A.S	N/A	N/A	Y	N	Y	N
(G) Metal Salt	N	N	N	N	N	N
(H) 1-Stage Lime	N	N	N	N	N	N
(I) 2-Stage Lime	N	N	N	N	N	N

Y - Can remove NH₄ or total N.

N - Need separate process or modification for NH₄ or total N removal.

N/A - Not applicable for effluent P limitation shown.

Table 2-9. Application Criteria Matrix - Effect of TBOD:TP Ratio < 20 on Process Applicability

Process	0.5 mg/l Effluent TP	1.0 mg/l Effluent TP	2.0 mg/l Effluent TP
	Application	Application	Application
(A) Phostrip	N/A	L	L
(B) Mod. Bardenpho	N/A	N	M
(C) A/O	N/A	N	M
(D) SBR	N/A	N	M
(E) UCT	N/A	N	M
(F) Mod. A.S.	N/A	N	M
(G) Metal Salt	L	L	L
(H) 1-Stage Lime	L	L	L
(I) 2-Stage Lime	L	L	L

L - No effect.

M - Marginal; may not meet effluent TP limitations without metal salt addition.

N - Cannot meet effluent phosphorus limits without metal salt addition.

N/A - Not applicable for effluent TP limitation shown.

Table 2-10. Application Criteria Matrix - Effect of Sludge Production on Process Applicability

Process	0.5 mg/l Effluent TP	1.0 mg/l Effluent TP	2.0 mg/l Effluent TP
	Application	Application	Application
(A) Phostrip	M	M	M
(B) Mod. Bardenpho	L	L	L
(C) A/O	L	L	L
(D) SBR	L	L	L
(E) UCT	L	L	L
(F) Mod. A.S.	L	L	L
(G) Metal Salt	H	M	M
(H) 1-Stage Lime	H	H	H
(I) 2-Stage Lime	H	H	H

L - Little or no increase in sludge production or handling (<30%).

M - Increased sludge production and/or handling (30 - 100%).

H - Substantial increase in sludge production and handling (> 100%).

Table 2-11. Application Criteria Matrix - Effect of O&M Requirements on Process Applicability¹

Process	0.5 mg/l Effluent TP	1.0 mg/l Effluent TP	2.0 mg/l Effluent TP
	Application	Application	Application
(A) Phostrip	N/A	M	M
(B) Mod. Bardenpho	N/A	M	L
(C) A/O	N/A	M	L
(D) SBR	N/A	M	L
(E) UCT	N/A	M	L
(F) Mod. A.S.	N/A	M	M
(G) Metal Salt	M	L	L
(H) 1-Stage Lime	S	M	M
(I) 2-Stage Lime	S	S	S

L - Little or no increase in O&M (0 - 10%).

M - Some increase in O&M (10 - 30%).

S - Substantial increase in O&M (> 30%).

N/A - Not applicable for effluent TP limitation shown.

¹ O&M - For process control, monitoring and operations excluding sludge handling.

CHAPTER 3

Phosphorus Removal by Biological Processes

3.1 Introduction and Theory

Conventional secondary biological treatment systems accomplish phosphorus removal by using phosphorus for biomass synthesis during BOD removal. Phosphorus is an important element in microorganisms for energy transfer and for such cell components as phospholipids, nucleotides, and nucleic acids. Attachment of a phosphate radical bond to adenosine triphosphate (ATP) results in the storage of energy (7.4 Kcal/mole P), which is available upon conversion of ATP back to adenosine diphosphate (ADP). Phosphorus is also contained in nucleotides such as nicotinamide adenine dinucleotide (NAD) and flavin adenine dinucleotide (FAD) which are used for hydrogen transfer during substrate oxidation-reduction reactions. Ribonucleic acid (RNA) and deoxyribonucleic acid (DNA) are composed of a deoxyribose sugar structure with attached amino acids of adenine, cytosine, guanine, and thymine or uracil. The deoxyribose molecules are attached by phosphorus bonds. Phosphorus may account for 10-12 percent of the RNA or DNA mass.

A typical phosphorus content of microbial solids is 1.5-2 percent based on dry weight. Wasting of excess biological solids with this phosphorus content may result in a total phosphorus removal of 10-30 percent, depending on the BOD-to-phosphorus ratio, the system sludge age, sludge handling techniques, and sidestream return flows.

In 1955, Greenburg *et al.* (1) proposed that activated sludge could take up phosphorus at a level beyond its normal microbial growth requirements. In 1959, Srinath (2) reported on batch experiments to conclude that vigorous aeration of activated sludge could cause the concentration of soluble phosphorus in mixed liquor to decrease rapidly to below 1 mg/l. In 1965, Levin and Shapiro (3) reported on enhanced biological phosphorus removal using activated sludge from the District of Columbia activated sludge plant. Over 80 percent phosphorus removal was observed by vigorous aeration of the sludge and without the addition of chemicals. They termed the high phosphorus removal "luxury uptake" by the microorganisms. In some experiments, a small amount of 2-4 di-nitrophenol was added that

inhibited phosphorus uptake, indicating the removal was of biological origin. They also observed volutin granules in the bacterial cells, which are reported in the microbial literature to contain polyphosphates. Acidification of the sludge resulted in the release of phosphorus, which led to a proposed treatment flow scheme of exposing return sludge to acidic conditions and stripping of phosphorus. Shapiro *et al.* (4) later observed high phosphorus uptake at the Baltimore sewage treatment plant and release in the bottom of the secondary clarifiers under conditions of zero or low dissolved oxygen (DO). They proposed that the return sludge could be intentionally exposed to such conditions prior to return to the aeration basin to strip out phosphorus. This work led to the development of the Phostrip process (5,6).

High levels of phosphorus removal were observed at various full-scale activated sludge plants in the United States, including the Rillings Road plant in San Antonio, Texas (7), the Hyperion plant in Los Angeles, California (8), and the Back River plant in Baltimore, Maryland (9). The three plants reported total phosphorus removals of 85-95 percent, and the phosphorus content of the waste sludge was 2-7.3 percent on a dry weight basis. All of the plants were of the plug flow configuration using diffused aeration, and the following operating characteristics were judged important in all or some of the plants to maximize phosphorus removal:

1. Require a DO concentration of 2.0 mg/l or greater from the middle to end of the plug flow aeration basins.
2. Prevent the recycle of phosphorus back to the activated sludge system via sludge handling streams.
3. Maintain aerobic conditions in the secondary clarifiers to prevent the release of phosphorus into the effluent.

Both Vacker (7) and Milbury (9) noted phosphorus release by the mixed liquor and an increase in soluble phosphorus concentration near the inlet of the activated sludge tanks at the Rillings Road and Back River plants, respectively. During the late 1960s to

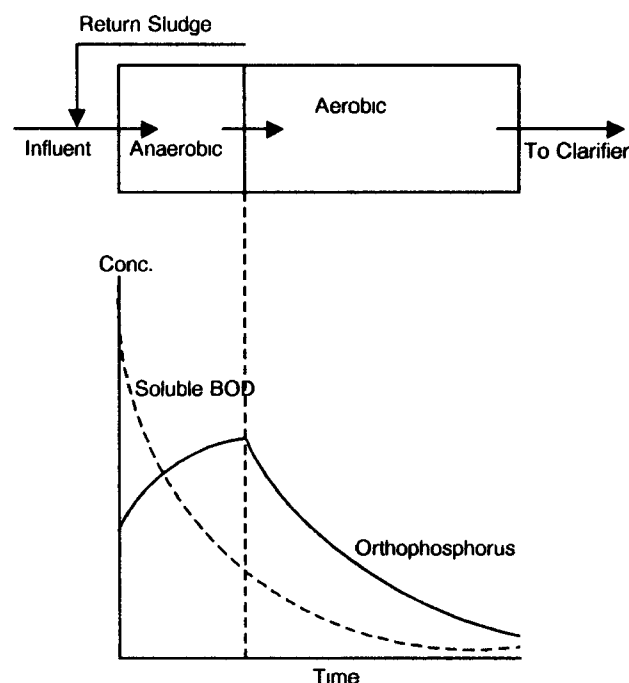
early 1970s there were varying opinions as to whether the excess phosphorus removal observed at these plants was due to chemical precipitation. In an attempt to provide a rational chemical removal explanation, one hypothesis was that the high pH, as a result of high aeration rates and carbon dioxide stripping at the end of the plug flow basins, encouraged the formation of a calcium hydroxyapatite precipitate (10).

In addition to the development of the Phostrip biological phosphorus removal process in the early 1970s in the United States, biological phosphorus removal was observed during the development of the Bardenpho four-stage biological nitrification-denitrification system by Barnard (11). The system consists of sequential anoxic-aerobic-anoxic-aerobic stages with an internal mixed liquor recycle from the first aerobic stage to the first anoxic stage. During a period of high phosphorus removal in a 100-m³/d (18-gpm) pilot-plant operation, Barnard observed a soluble phosphorus concentration of 0.3 mg/l in the final aerobic basin. He recognized that phosphorus was being released in the designated second "anoxic" basin, which was actually experiencing anaerobic conditions (absence of both nitrate nitrogen and DO) and that it was being taken up in the final aerobic stage. This led him to conclude that biological phosphorus removal was possible in activated sludge systems provided that an aerobic stage was preceded by an anaerobic stage where phosphorus release occurred. It was also noted that when a high level of phosphorus removal was reported in plug flow U.S. plants, phosphorus release occurred near the inlet of the aeration basin followed by phosphorus uptake along the length of the basin where the DO concentration increased.

In a later paper, Barnard (12) proposed the use of a separate anaerobic basin ahead of the Bardenpho nitrogen removal system or ahead of aerobic basins when nitrogen removal was not necessary. The former was called the Modified Bardenpho process and the latter the Phoredox process. Phoredox was derived from "phosphorus" and "redox potential," which is at a lower level in the anaerobic phosphorus release zone. Figure 3-1 shows phosphorus release and uptake characteristics of such biological phosphorus removal systems that employ sequential anaerobic-aerobic contacting.

Following Barnard's pilot-plant work, full-scale plants were modified at Johannesburg, South Africa, to investigate the feasibility of biological phosphorus removal. At the Alexander plant, surface aerators near the inlet of an activated sludge basin were turned off to create an anaerobic zone (13). An overall nitrogen removal of 85 percent and a total phosphorus removal of 46 percent were reported. At the Olifantsvlei plant, various combinations of surface aerators were turned off in the four-stage system

Figure 3-1. Biological phosphorus and BOD removal due to anaerobic-aerobic contacting.



and an effluent soluble phosphorus concentration of 0.9 mg/l was reported (14). Based on this work, a modified Bardenpho system was designed for a 150,000-m³/d (39-mgd) facility at the Johannesburg Goudkoppies wastewater plant that became operational in 1978 (15). In the late 1970s, a modified Bardenpho plant was started up at Palmetto, Florida (16), and a portion of the Largo, Florida facility was converted to the A/O process (17), an anaerobic-aerobic biological phosphorus removal system to be discussed later.

3.1.1 Biological Phosphorus Removal Mechanism

The generally accepted theory for biological phosphorus removal is that anaerobic-aerobic contacting results in a competitive substrate utilization and selection of phosphorus-storing microorganisms (18,19). An understanding of the steps involved in the biological phosphorus removal mechanism provides a useful insight into the factors that can affect the performance of biological phosphorus removal systems. The following observations by various investigators are presented as a background to the proposed mechanism.

Funs and Chen (20) examined activated sludge from the Baltimore Back River and the Seneca Falls, New York treatment plants when the plants were exhibiting high levels of phosphorus removal. They concluded that the organism associated with phosphorus removal belonged to the *Acinetobacter* genus. These bacteria are short, plump, gram-negative rods with a size of 1-1.5 μ m. They appear in pairs, short chains, or clusters. They also subjected a pure culture to

batch anaerobic-aerobic cycles and noted excess phosphorus removal when acetate was fed to the system. They postulated that the anaerobic phase in excess phosphorus removal systems was important for the production of simple carbohydrates such as ethanol, acetate, and succinate, which serve as carbon sources for *Acinetobacter*. Contrary to later findings, they felt that the simple carbohydrates were assimilated by the *Acinetobacter* in the aerobic phase of the cycle. Fuhs and Chen also found that a significant phosphorus release rate could be promoted by the addition of carbon dioxide during the anaerobic phase, which also lowered the pH. This was also observed by Deinema (21).

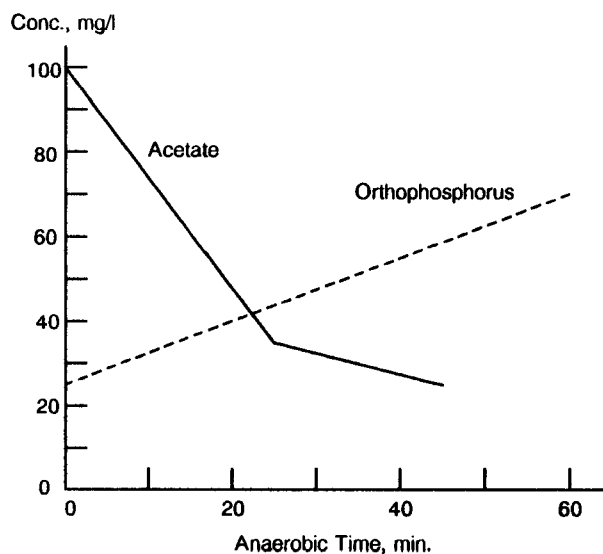
Other investigators also reported observing significant levels of *Acinetobacter* in biological excess phosphorus removal systems (22-24). Lotter (24) also found significant levels of *Aeromonas* and *Pseudomonas* which are capable of polyphosphate accumulation. Hascoet *et al.* (25) also noted the presence of *Bacillus cereus* in addition to *Acinetobacter* and Suresh *et al.* (26) found small amounts of *Pseudomonas vesiculcris*, besides *Acinetobacter*, in samples cultured from an anaerobic-aerobic phosphorus removing pilot plant. Brodich (27) noted that the removal of phosphorus in a system containing *Acinetobacter* became significant only after the development of an *Aeromonas* population. He postulated that the *Aeromonas* bacteria served the important function of producing fermentation products in the anaerobic phase for the *Acinetobacter*. Lotter and Murphy (28) noted an increase of *Pseudomonas* and *Aeromonas* in biological phosphorus removal systems. They also found that these species of bacteria and a species of *Acinetobacter* accomplished denitrification in anoxic zones of biological nitrogen removal systems. Osborn and Nicholls (15) reported rapid biological phosphorus uptake during nitrate reduction in the absence of DO, indicating that phosphorus uptake may be occurring with denitrifying bacteria. Hascoet (25) also reported phosphorus release in anoxic zones by *Acinetobacter* provided that there was a relatively high level of substrate availability.

Various investigators have observed a decrease in soluble substrate and an increase in orthophosphate concentrations in the anaerobic zone of anaerobic-aerobic sequenced biological phosphorus removal systems. Hong (29) showed a soluble BOD₅ (SBOD) concentration decrease from 45 to 15 mg/l and an orthophosphorus concentration increase from 6 to 24 mg/l in the anaerobic zone. Ekama *et al.* (30) related phosphorus release in the anaerobic zone to the presence of a soluble, readily biodegradable substrate. The concentration of a soluble readily biodegradable substrate can be determined from the increase in the oxygen uptake rate measurements of a batch activated sludge sample after the addition of influent. Rensink's (31) work on *Acinetobacter* led

him to investigate the change in acetate and soluble phosphorus concentrations in the anaerobic phase.

Figure 3-2 shows the decrease in acetate concentration and increase in orthophosphate concentration as a function of the anaerobic time. The molar ratio of acetate utilization to phosphorus release was 1.3.

Figure 3-2. Acetate assimilation and phosphorus release vs. anaerobic time (31).



Fukase (32), using fill-and-draw reactors, observed an acetate utilization to phosphorus release molar ratio of 1.0. Arvin (33) reported 0.7, Rabinowitz (34) 0.6, and Wentzel (35) 1.0 from batch studies using sludge from excess biological phosphorus removal systems. Rabinowitz (34) also found that the phosphorus release magnitude and rate were affected by the type of substrate. The amount of phosphorus release for each substrate in decreasing order was sodium acetate, propionic acid, glucose, acetic acid, and butyric acid. On the other hand, Jones (36) observed a greater phosphorus release in declining order of butyric acid, ethanol, acetic acid, methanol, and sodium acetate.

Release and uptake of metal ions have been observed during phosphorus release in biological phosphorus removal systems. A summary of the data has been presented by Comeau (37) and is shown in Table 3-1.

The understanding of the biological phosphorus removal mechanism was significantly advanced with the observations on storage of carbohydrate products within biological cells in the anaerobic zone and phosphorus-containing volutin granules in the aerobic zone. The most commonly reported

Table 3-1. Molar Ratios of Ions Co-Transported with Phosphorus.

	Ref. 38	Ref. 39	Ref. 37	Ref. 37	Ref. 37
Mg ²⁺ /P	0.26	0.32	0.24	0.28	0.27
K ⁺ /P	0.27	0.23	0.34	0.20	0.23
Ca ²⁺ /P	0.00	-	0.06	0.09	0.12
Na ⁺ /P	-	-	0.00	0.00	0.00
charges/P	0.79	0.87	0.94	0.94	1.01

anaerobic intracellular storage product has been polyhydroxybutyrate (PHB).

PHB has been found in biologically-removed phosphorus sludges by Timmerman (40) and in *Acinetobacter* by Nicholls and Osborn (41). Lawson and Tonhazy (23) isolated *Acinetobacter* and showed that these bacteria could accumulate PHB and polyphosphates. Deinema (42) also observed PHB in a strain of phosphorus-removing *Acinetobacter*. Senior (43) hypothesized that certain bacteria will accumulate PHB during temporary deprivation of oxygen. Buchan (22) reported that PHB increased in bacterial cells while polyphosphate granules decreased in size or disappeared in the anaerobic zone of biological phosphorus removal systems.

PHB synthesis and degradation are described by Gaudy (44). PHB is formed in the cell under anaerobic conditions from acetoacetate serving as a hydrogen acceptor. Acetate entering the bacterial cells under anaerobic conditions can be converted to acetyl-COA provided energy is available, and acetyl-COA can be converted to acetoacetate since the cell has a limited supply of the enzyme COA. During oxidative conditions, PHB is oxidized to acetyl-COA, which enters the Krebs cycle. PHB oxidation does not occur until nearly all of the endogenous carbon is used up according to Gaudy.

As described previously, Levin (3) reported finding volutin granules in sludge samples during biological phosphorus removal. In an extensive evaluation of biological phosphorus removal, Harold (45) stated that phosphorus was likely stored as polyphosphates within volutin granules. Volutin granules contain lipids, protein, RNA, and magnesium in addition to polyphosphates. The granules are visible under the light microscope and can also be identified by staining with either toluidine dye, which results in a reddish-purple color, or with a methylene blue technique, which results in a dark purple color. A high electron beam directed on the microorganisms will also volatilize the polyphosphate contained in the volutin granules leaving holes in the cells.

Sell (46) photographed large masses of polyphosphate granules contained in phosphorus removing sludge during a cold temperature laboratory

investigation of biological phosphorus removal. Buchan (22) analyzed the biological species obtained from aerobic zones of various South African activated sludge plants accomplishing biological phosphorus removal. His analysis showed that the intracellular polyphosphate granules contained an excess of 25 percent phosphorus. In the anaerobic zone, the large polyphosphate granules had dispersed into smaller granules and some cells had released virtually all of their accumulated phosphorus.

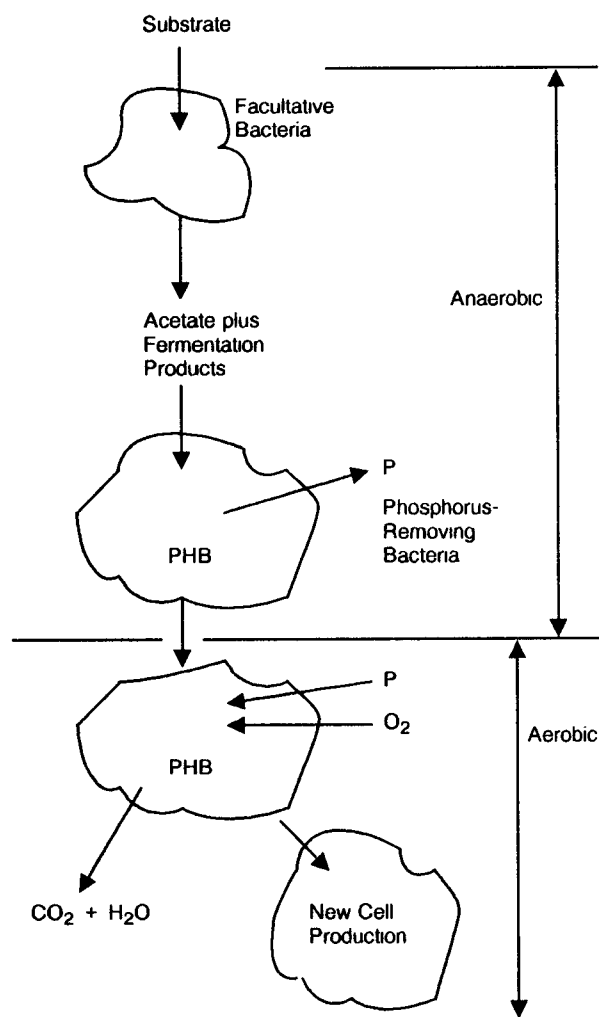
The proposed biological phosphorus removal mechanism (18,19) is summarized in Figure 3-3. Acetate and other fermentation products are produced from fermentation reactions by normally-occurring facultative organisms in the anaerobic zone. A generally accepted concept is that these fermentation products are derived from the soluble portion of the influent BOD and that there is not sufficient time for the hydrolysis and conversion of the influent particulate BOD. The fermentation products are preferred and readily assimilated and stored by the microorganisms capable of excess biological phosphorus removal. This assimilation and storage is aided by the energy made available from the hydrolysis of the stored polyphosphates during the anaerobic period. The stored polyphosphate provides energy for active transport of substrate and for formation of acetoacetate, which is converted to PHB. The fact that phosphorus-removing microorganisms can assimilate the fermentation products in the anaerobic phase means that they have a competitive advantage compared to other normally-occurring microorganisms in activated sludge systems. Thus, the anaerobic phase results in a population selection and development of phosphorus-storing microorganisms. Rensink (31) has pointed out that *Acinetobacter* are relatively slow growing bacteria and that they prefer simple carbohydrate substrates. Thus, without the anaerobic phase, they may not be present at significant levels in conventional activated sludge systems.

During the aerobic phase, the stored substrate products are depleted (22) and soluble phosphorus is taken up, with excess amounts stored as polyphosphates. An increase in the population of phosphorus-storing bacteria is also expected as a result of substrate utilization. The above mechanism indicates that the level of biological phosphorus removal achieved is directly related to the amount of substrate that can be fermented by normally-occurring microorganisms in the anaerobic phase and subsequently assimilated and stored as fermentation products by phosphorus-removing microorganisms, also in the anaerobic phase.

3.2 Applications

The recent developments leading to a better understanding of the conditions causing excess

Figure 3-3. Schematic of biological phosphorus removal mechanism.



biological phosphorus removal help to explain the earlier observations on excess phosphorus removal reported for full-scale facilities. It is apparent now that sufficient BOD was present and oxygen was limiting so that fermentation conditions likely occurred at the front end of the relatively long, narrow aeration basins of these plants. Since these observations, three major proprietary biological phosphorus removal processes that employ more definitive anaerobic fermentation zones have been commercialized. These processes are, in order of development, the Phostrip process, the modified Bardenpho process, and the A/O process. These processes as well as other non-proprietary systems will be described in this section. Other options used are the UCT process, sequencing batch reactors (SBRs), and operationally modified activated sludge systems.

3.2.1 Process Descriptions

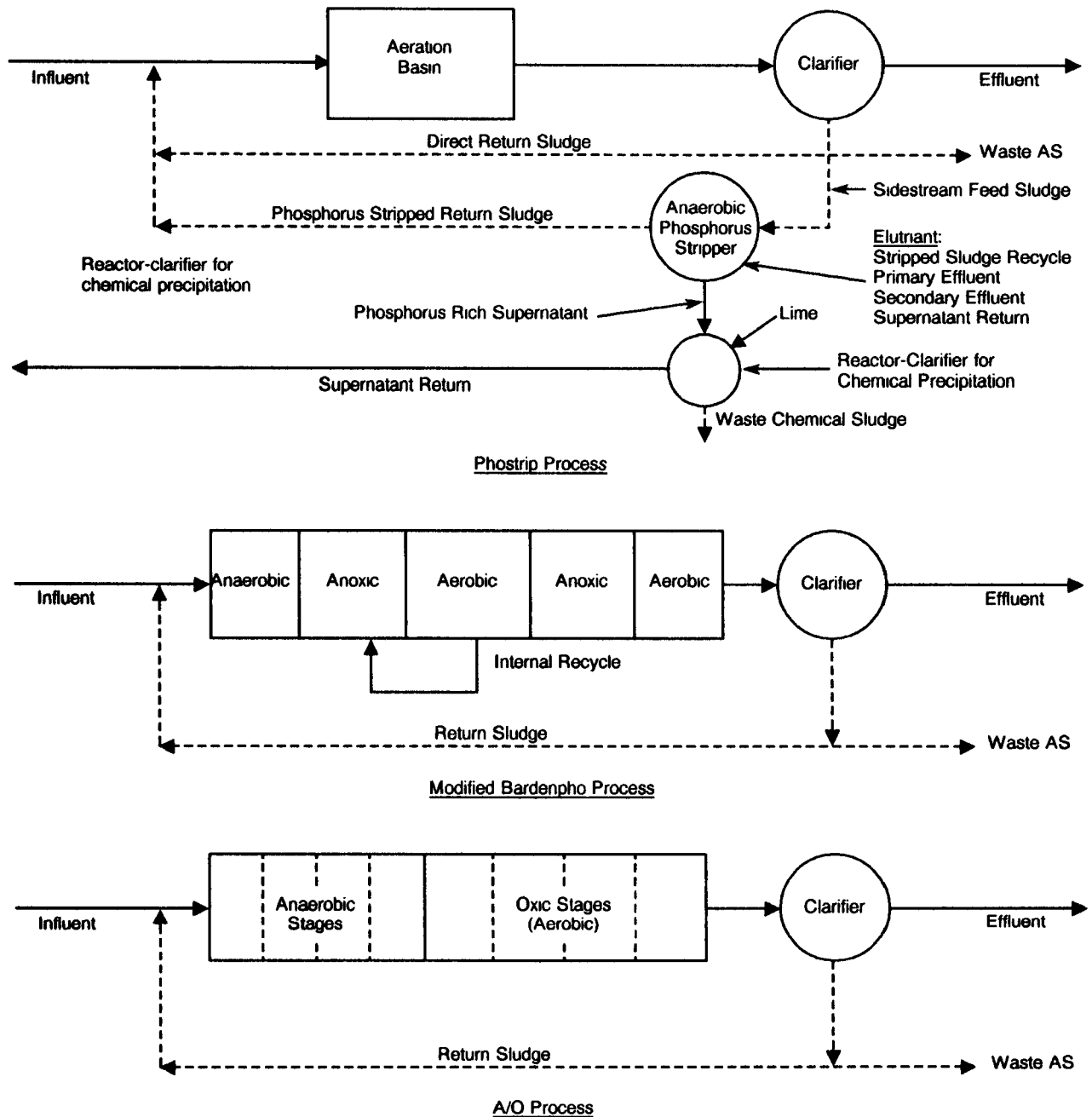
The three commercial biological phosphorus removal processes are shown in Figure 3-4. The Phostrip

process was first proposed by Levin in 1965 (3). Pilot plant data were collected at a number of municipal plants from 1970 to 1973 and demonstrated high levels of phosphorus removal. In 1973, the Seneca Falls, New York activated sludge plant was converted to the Phostrip process and evaluated (6). The process combines both biological and chemical phosphorus removal and has been referred to as a sidestream process since a portion of the return activated sludge flow is diverted for phosphorus stripping and subsequent precipitation with lime. An advantage of the Phostrip process is that an effluent concentration of less than 1 mg/l total phosphorus can be achieved with less dependence on the BOD strength of the influent wastewater. A large percentage of the phosphorus removed is tied up as a lime sludge, which causes less concern than handling a phosphorus-rich waste biological sludge. Compared to chemical addition to an activated sludge aeration basin for phosphorus precipitation, the Phostrip process may require a lower chemical dosage and cost, since the lime dosage is a function of the alkalinity and not the amount of phosphorus to be removed, as is the case for alum and iron salts. This potential advantage is a function of wastewater alkalinity, phosphorus concentration, and relative chemical costs. The process may require more operator skill and control relative to the stripper tank operation and lime feeding. In addition, as discussed in Chapter 4, significant problems have been reported for lime storage and handling systems.

The sidestream flow diverted to the anaerobic phosphorus stripper tank is normally 10-30 percent of the influent flow. The stripper tank also functions as a gravity sludge thickener. The average solids detention time (SDT) in the stripper tank can be 5-20 hours, with 8-12 hours being typical (47,48). The SDT equals the mass of solids in the sludge blanket divided by the mass of solids removed per day in the tank underflow. Soluble phosphorus is released in the anaerobic stripper tank. It is not known if the phosphorus release is due to the exact same mechanism as described for the anaerobic-aerobic activated sludge contacting sequence. Fermentation products for the biological phosphorus-removing organisms may be derived from the metabolism of hydrolyzed solids and from organics released from lysed bacteria in the stripper. The released soluble phosphorus may be from biological phosphorus-removing microorganisms and from lysed bacteria. Soluble phosphorus is transferred to the supernatant either by recycling the stripper underflow to the stripper influent or by passing an elutriation stream through the stripper. The elutriation stream may be primary effluent, secondary effluent, or supernatant from the lime precipitation reactor.

The overflow from the stripper tank is continuously fed to the chemical treatment tank where lime is added for phosphorus removal. Two approaches have

Figure 3-4. Commercial biological phosphorus removal processes.



been proposed for the precipitation and removal of the chemical sludge. The first, as shown in Figure 3-4, is to use a separate reactor-clarifier unit for treatment of the stripper overflow. The second is to add the lime to the overflow but to settle the chemical precipitate in the primary clarifier. The separate treatment is the more common approach. The underflow solids from the stripper tank are returned to the aeration tank where biological uptake of phosphorus occurs. Control of the sidestream feed rate to the stripper tank affects the distribution

between phosphorus removal by chemical precipitation or in the waste biological sludge.

A summary of typical recommended design criteria for the Phostrip, Modified Bardenpho, and A/O processes is shown in Table 3-2. A significant design feature for the three processes is the operating organic loading. The Phostrip process is not confined to a narrow range of loadings as are the other two processes and has been recommended for a wide range of activated sludge operations. This is

Table 3-2. Typical Operating Conditions for Biological Phosphorus Removal Processes (22,44,49,50).¹

Phostrip		Modified Bardenpho		A/O		A/O plus Nitrification	
Parameter	Value	Parameter	Value	Parameter	Value	Parameter	Value
AS System							
F/M, kg TBOD/ kg MLVSS/d	--1	F/M, kg TBOD/ kg MLVSS/d	0.1-0.2	F/M, kg TBOD/ kg MLVSS/d	0.2-0.7	F/M, kg TBOD/ kg MLVSS/d	0.15-0.25
SRT, days ²	--1	SRT, days ²	10-30	SRT, days ²	2-6	SRT, days ²	4-8
MLSS, mg/l	600-5,000	MLSS, mg/l	2,000-4,000	MLSS, mg/l	2,000-4,000	MLSS, mg/l	3,000-5,000
HRT, hr ³	1-10	HRT, hr ³		HRT, hr ³		HRT, hr ³	
		Anaerobic	1-2	Anaerobic	0.5-1.5	Anaerobic	0.5-1.5
		Anoxic 1	2-4	Aerobic	1-3	Anoxic	0.5-1.0
		Nitrification (Aerobic 1)	4-12			Nitrification	3.5-6.0
		Anoxic 2	2-4				
		Aerobic 2	0.5-1.0				
Phostrip Stripper							
Feed, % of inf. flow	20-30	Return Sludge, % of inf. flow	100	Return Sludge, % of inf. flow	25-40	Return Sludge, % of inf. flow	20-50
SDT, hr	5-20	Int. Recycle, % of inf. flow	400			Int. Recycle, % of inf. flow	100-300
Sidewater Depth, m	6.1						
Elutriation Flow, % of stripper feed flow	50-100						
Underflow, % of inf. flow	10-20						
P Release, g P/g VSS	0.005-0.02						
Reactor-Clarifier							
Overflow Rate, m ³ /m ² /d	48						
pH	9-9.5						
Lime Dosage, mg/l	100-300						

¹ Based on activated sludge system design.

² Average mass of solids in the system divided by average mass of solids wasted daily.

³ Hydraulic retention time, volume divided by influent flow rate.

due to the fact that Phostrip performance is related more to the stripper operation and chemical treatment step. The A/O process is generally designed as a high-rate activated sludge system, while the Modified Bardenpho process is generally designed at relatively low overall loading rates due to the detention time required for nitrification and denitrification.

The Modified Bardenpho process, marketed by the Eimco Process Equipment Company of Salt Lake City, Utah, is both a nitrogen and a phosphorus removal system. As Figure 3-4 illustrates, the influent and return sludge are contacted in an anaerobic tank to promote fermentation reactions and phosphorus release prior to passing the mixed liquor through the four-stage Bardenpho system. The original development of the four-stage Bardenpho

process was described by Barnard (50) to provide more than 90 percent nitrogen removal without using an exogenous carbon source. In the first anoxic stage, nitrate nitrogen contained in the internal recycle from the nitrification stage is reduced to nitrogen gas (denitrification) by metabolizing influent BOD using nitrate oxygen instead of DO. About 70 percent of the nitrate nitrogen produced in the system is removed in the first anoxic stage. In the nitrification (first aerobic) stage, BOD removal, ammonium nitrogen oxidation, and phosphorus uptake occurs. The second anoxic stage provides sufficient detention time for additional denitrification by mixed liquor endogenous respiration, again using nitrate oxygen instead of DO. The final aerobic stage provides a short period of mixed liquor aeration prior to clarification to minimize anaerobic conditions and phosphorus release in the secondary clarifier.

The resultant Modified Bardenpho design solids retention time (SRT), based on the solids inventory in all the aerobic and anoxic stages, is typically 10-20 days depending on the wastewater temperature and influent nitrogen concentration. In some designs, the tank volumes have been increased above the nitrogen removal requirements to provide an extended SRT of 20-30 days for the purpose of sludge stabilization. In this way, further sludge digestion is not included in the facility design. As will be described in Section 3.3, this design approach results in less sludge production and phosphorus removal per unit of influent BOD removed.

The A/O process shown in Figure 3-4 is marketed in the United States by Air Products and Chemicals, Inc. (17) and is similar to the Phoredox concept described by Barnard (12), except that the anaerobic and aerobic stages are divided into a number of equal size complete mix compartments. Typically, three compartments have been used for the anaerobic stage and four for the aerobic stage. The key features of the A/O process are its relatively short design SRT and high design organic loading rates. Compared to the Modified Bardenpho process, this results in greater sludge production and more phosphorus removal per unit of BOD removed in the system. However, the use of further sludge stabilization methods, such as anaerobic or aerobic digestion, must consider the amount of phosphorus released during stabilization and the effect of recycle streams from the stabilization units on facility performance.

As shown in Table 3-2, the A/O process can also be used where nitrification and/or denitrification are required. The modified flow scheme incorporates an anoxic stage for denitrification between the anaerobic and aerobic stages and is called the A²/O process. The anoxic stage is also divided into three equal-size, complete mix compartments. Mixed liquor is recycled from the end of the nitrification stage to feed nitrate nitrogen into the anoxic stage for denitrification. Internal recycle flows of 100-300 percent have been used. Nitrate nitrogen removals of 40-70 percent can be accomplished this way.

The use of SBR systems for secondary treatment has gained increased popularity in the United States during the late 1970s and early 1980s. An evaluation of SBR treatment capabilities, design aspects, full-scale installations, and advantages has been documented for conventional activated sludge treatment applications (51). Though not a new treatment concept, with reported operations dating back to the early 1900s, the recent surge of interest has been related to new and improved hardware devices and to the successful EPA-funded, full-scale, 20-month demonstration and evaluation of a 1,330-m³/d (0.35-mgd) facility at Culver, Indiana (52). Unique hardware for the system consists of motorized or pneumatically-actuated valves, level

sensors, automatic timers, microprocessor controllers, and effluent withdrawal decanters. The SBR treatment concept and operational flexibility makes it an obvious candidate for employing anaerobic-aerobic contacting for biological phosphorus removal. Biological phosphorus removal was demonstrated in the full-scale Culver, Indiana facility during June and July 1984 (53).

A schematic of an SBR operation for biological phosphorus removal is shown in Figure 3-5. The SBR system is a fill-and-draw activated sludge system. A single tank provides for activated sludge aeration, settling, effluent withdrawal, and sludge recycle. Biological phosphorus removal was accomplished in two SBR basins at the Culver facility that were operating at substantially different average food-to-microorganism (F/M) loadings of 0.16 and 0.42 kg total BOD₅ (TBOD)/kg MLVSS/d, respectively. The operation steps consist first of a fill period where flow is diverted to one of the SBR tanks while the other tank(s) operates in the reaction, settle, effluent withdrawal, or idle operation sequences. After the fill period, the reactor contents are mixed but not aerated to provide the anaerobic fermentation period for phosphorus release and uptake of soluble fermentation products. The next step is the react or aeration period followed by a selected settling time when both aeration and mixing are stopped. The effluent is then withdrawn and, depending on the influent flow rate, a variable length idle time may occur. The operating times for this biological phosphorus removal sequence in the two differently-loaded SBR basins at Culver, Indiana, are shown in Table 3-3 (53).

Figure 3-6 shows a further modification of the Modified Bardenpho process. This modification was developed at the University of Capetown in South Africa (30) and has been termed the UCT process. As shown, the return activated sludge is directed to the anoxic stage instead of the anaerobic stage as in the Modified Bardenpho process. The basis for this development was previous work with biological phosphorus removal systems that indicated initial phosphorus removal efficiency could be negatively affected by nitrate nitrogen entering the anaerobic stage (16,54,55). Nitrate will serve as an electron acceptor during the biological oxidation of BOD entering the anaerobic stage. This results in competition for the soluble, readily biodegradable BOD that would normally be converted to fermentation products for use by the biological phosphorus-removing bacteria in the anaerobic zone in the absence of nitrate nitrogen. The relative ratio between the nitrate nitrogen in the sludge recycled to the anaerobic stage in a Modified Bardenpho or A/O process and the available, readily degradable soluble BOD in the influent to that zone will determine if sufficient BOD will remain after denitrification reactions occur to produce a necessary level of

Figure 3-5. Biological phosphorus removal using a Sequencing Batch Reactor.

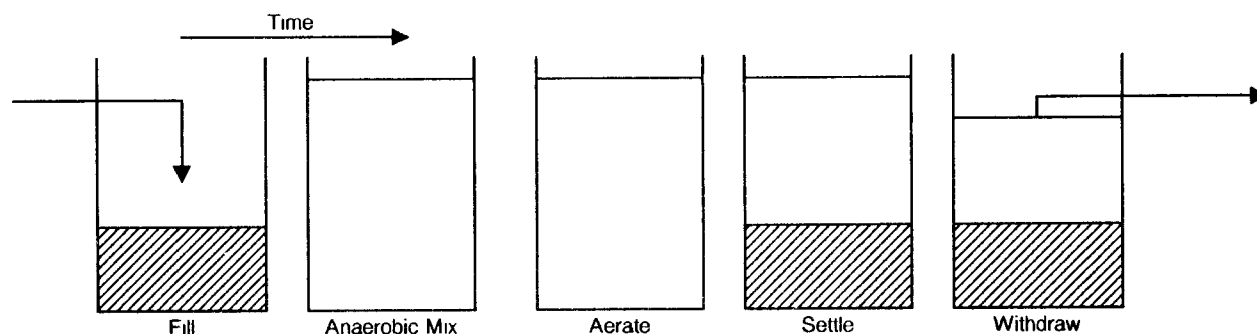


Table 3-3. SBR Operating Sequence - Biological Phosphorus Removal.

Period	Low Loaded	High Loaded
	hr	hr
Fill and Anaerobic Mix	1.8	3.0
Aerate	1.0	0.4
Settle	1.0	0.7
Withdraw	0.4	0.7
Idle	0.6	0.0

fermentation products for biological phosphorus removal. For wastewaters with a relatively high TKN:BOD ratio, the effect of nitrate nitrogen in the return sludge on anaerobic zone fermentation may be significant for these two processes.

In contrast, the anoxic stage of the UCT process is designed and operated to produce a very low nitrate nitrogen concentration. The recycle of mixed liquor from the anoxic stage to the anaerobic stage thereby provides optimum conditions for conversion of available soluble BOD to fermentation products. The mixed liquor recycle from the aerobic stage to the anoxic stage (recycle 2) can be controlled to assure a minimal nitrate nitrogen concentration in recycle 1, while achieving some level of nitrogen removal in the anoxic zone. The process has generally been recommended for wastewaters with influent TKN:COD ratios of greater than 0.08 or influent COD:TKN ratios of less than 12.0 for South African applications. Gerber *et al.* (56) compared UCT and Modified Bardenpho process performance in pilot-scale studies. At a COD:TKN ratio of 9.5, he found no performance difference.

A modified UCT process is also shown in Figure 3-6. In this case, the first anoxic zone is designed to reduce only the nitrate nitrogen in the return activated sludge. The second anoxic zone is designed for a much higher quantity of nitrate nitrogen removal as mixed liquor is recycled to it from the nitrification zone.

Another approach to accomplish biological phosphorus removal is to make operational changes in existing activated sludge systems to create an anaerobic fermentation zone ahead of the aeration zone. Figure 3-7 indicates this approach. In practice, it typically involves turning off air flow or aerators in the front of the activated sludge basin. As described in Section 3.1, this technique was demonstrated during the earlier investigations of phosphorus removal with the Bardenpho process. The plug flow plants in the United States, for which high levels of phosphorus removal were reported, likely had insufficient aeration at the front end of the aeration basins and inadvertently promoted the anaerobic-aerobic contacting sequence. Full-scale U.S. plant operation modifications that have been shown to accomplish biological phosphorus removal are the 18,130-m³/d (4.8-mgd) Walt Disney World resort complex wastewater treatment facility in Orlando, Florida, and the 13,250-m³/d (3.5-mgd) DePere, Wisconsin wastewater treatment facility (47). Both plants had about half of their original aeration volume converted to non-aerated zones.

In addition to the designs presented, other modifications have been proposed that combine chemical treatment with anaerobic-aerobic staged activated sludge systems. Alum can be added to biological phosphorus removal systems as a polishing step to reduce the total phosphorus concentration to less than 1 mg/l where insufficient biological phosphorus removal occurs. Alum is added to the mixed liquor prior to the secondary clarifier at the Palmetto Bardenpho facility (49). The addition results in an effluent total phosphorus concentration of less than 1 mg/l vs. 2-3 mg/l when no alum is added.

Figure 3-8 shows a combination biological anaerobic-aerobic system used by Rensink (54) that also incorporates a stripper for phosphorus removal. The stripper consisted of a complete mix tank for anaerobic contacting of a sidestream of return activated sludge followed by a clarifier for separation of the stripped sludge. This combination achieved

Figure 3-6. UCT process flow schematics.

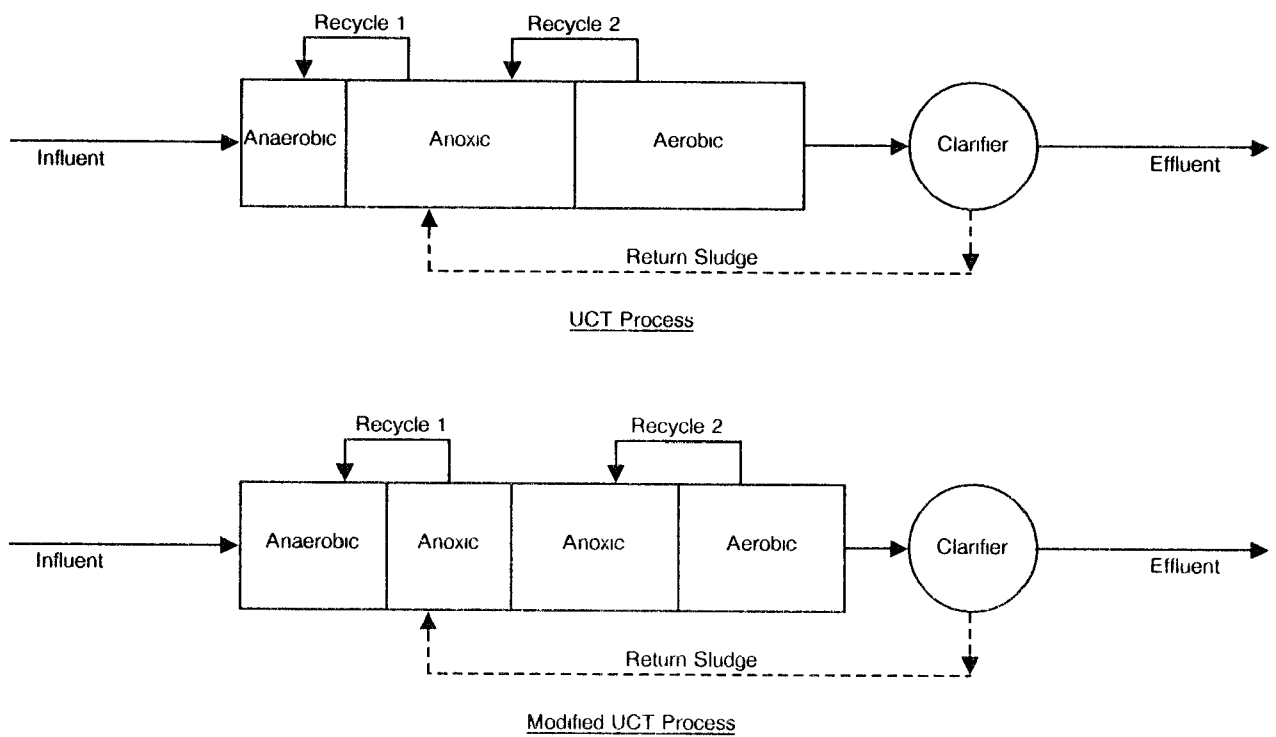


Figure 3-7. Operationally modified activated sludge system for biological phosphorus removal.

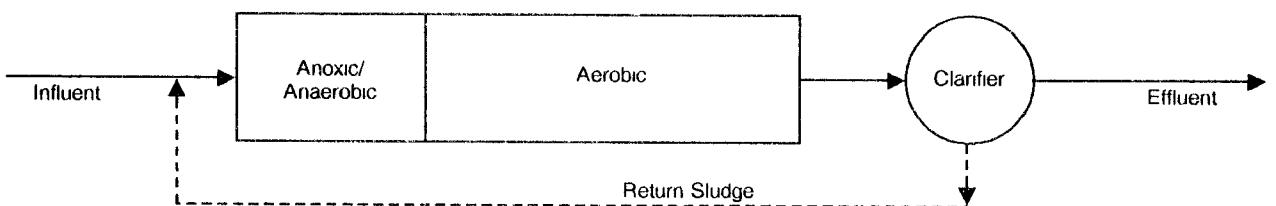
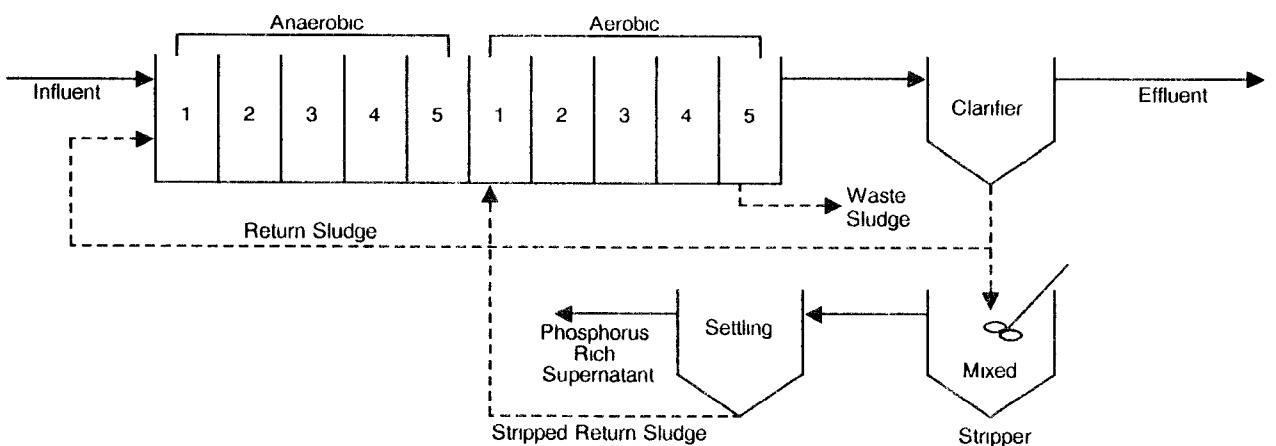


Figure 3-8. Combination biological phosphorus removal system.



more than 97 percent total phosphorus removal compared to 40-50 percent removal for the anaerobic-aerobic sequence without the stripper. The activated sludge system was operated with a relatively low organic loading and nitrification was also occurring. High nitrate production may have affected the phosphorus removal efficiency of the anaerobic-aerobic system without the stripper.

3.3 Performance

An inventory of full-scale biological phosphorus removal facilities was identified as of April 1984 by Tetreault *et al.* (47). Thirty biological phosphorus removal facilities were identified as being in operation, construction, or design with 28 of these being either a Phostrip, Modified Bardenpho, or A/O process. At that time, 11 of the facilities were in operation with five being Phostrip installations. Three were modified Bardenpho installations, two were operationally modified activated sludge systems, and one was an A/O system.

3.3.1 Phostrip Performance

Table 3-4 summarizes the basic design information for full-scale Phostrip plants, and Table 3-5 provides performance data summaries for phosphorus removal by the Phostrip plants. The Seneca Falls, New York plant was a full-scale demonstration project, and the performance data are for one month of intensive plant monitoring. The Phostrip process is used in the first-stage activated sludge system of a two-stage nitrification system at the Lansdale, Pennsylvania plant. The second-stage nitrification system consists of trickling filters and clarification. Primary treatment is not provided at this plant, but a 24-hour detention, in-line flow equalization basin is used. For the June to August 1984 test period, the Lansdale effluent contained an average of 0.8 mg/l total phosphorus. This performance illustrates the ability of the Phostrip process to produce low effluent phosphorus concentrations with weak wastewaters, as the influent TBOD averaged only 41 mg/l. Reactor clarifier overflow from the lime treatment step and secondary effluent were used as the stripper tank elutriant sources. The secondary effluent was nitrified and the stripper overflow contained up to 3.0 mg/l nitrate nitrogen (47). The orthophosphorus concentration in the stripper overflow averaged 20.0 mg/l, and the stripper SDT was 20 hours. Lime scaling occurred in the reactor-clarifier elutriation return line.

The effluent phosphorus concentrations shown for Adrian, Michigan, are for samples after the second-stage nitrification step and filtration. A first-stage activated sludge system incorporating the Phostrip process is preceded by primary treatment and in-line equalization. For the performance data shown in Table 3-5, the plant influent TBOD averaged only 78 mg/l. The upset problems noted were associated with

mechanical problems with the anaerobic digesters. During these periods sludge was temporarily stored in the treatment system and the stripper tank was overloaded.

The Savage, Maryland plant (also referred to as the Little Patuxent plant) is a two-stage nitrification activated sludge system following primary clarification. The Phostrip process is operated within the high-rate first-stage activated sludge system. Partial nitrification occurs in this stage. The nitrification stage is followed by filtration and chemical polishing to consistently produce effluent total phosphorus concentrations below 1.0 mg/l. The effluent phosphorus concentration values shown in Table 3-5 are for samples taken after the first-stage activated sludge system. The improved effluent phosphorus concentration for April 1985 reflects operating changes made at the plant. These consisted of changing the first-stage activated sludge system from a step feed to a plug flow mode and reducing the reactor-clarifier elutriation overflow rate to the stripper tank. Stripper underflow was used to supplement the reduced reactor-clarifier elutriation flow. The reduced reactor-clarifier elutriant flow rate was thought to be responsible for improved stripper performance, resulting in an overflow orthophosphorus concentration of 17.6 mg/l in April 1985 compared to 7.2 mg/l for the July 1984 operation.

The Southtowns, New York facility has a relatively low influent phosphorus concentration, and the Phostrip system is reportedly not operated continuously when the effluent phosphorus limit is met by secondary treatment and filtration alone (59). The Amherst, New York facility is not operating its Phostrip system, as two-point ferric chloride addition has been found to effectively treat a presently lower influent total phosphorus concentration of 3-4 mg/l. Prior to this the plant experienced a number of mechanical and operating problems. Mechanical problems included scaling in the lime feed line, malfunction of lime feed pumps, and freezing in the lime slurry line. Operational problems were encountered when the activated sludge system MLSS concentration was increased to provide extended aeration conditions, which resulted in overloading the stripper tank.

The full-scale Reno-Sparks, Nevada facility used the Phostrip process for phosphorus removal as part of a plant expansion design after successful results with a 23,000-m³/d (6-mgd) plant-scale evaluation (61). The 113,700-m³/d (30-mgd) expanded facility employs five parallel anaerobic stripper units. Elutriation is accomplished by recycling sludge from the bottom of the stripper to the stripper inlet to release phosphorus into the stripper supernatant. During portions of the initial startup phase, the effluent total phosphorus exceeded 1-3 mg/l. This was attributed to colloidal carry-over of lime-

Table 3-4. Basic Design Information for Full-Scale Phostrip Plants (57).

	Seneca Falls, NY	Landsdale, PA	Adrian, MI	Savage, MD	Southtowns, NY	Amherst, NY	Reno- Sparks, NV
Design flow, m ³ /d	3,400	9,500	26,500	56,800	60,600	90,900	113,700
Final eff. TP std., mg/l	1.0	2.0	1.0	0.3 ¹	1.0	1.0	0.5 ¹
Aeration by <u>O</u> xxygen or <u>A</u> ir	A	A	A	A	O	O	A
Aeration mode	Complete Mix	Plug Flow	Conv.	Plug Flow or Step Feed	Plug Flow	High Rate	Plug Flow
Nitrification, 1- or 2- stage sec. treatment	1	2	2	2	1	2	1
Equalization	No	Yes	Yes	Yes	No	Yes	No
Final filtration	No	No	Yes	Yes	Yes	Yes	Planned
Sludge handling	Thickening, Anaer. Dig.	Thickening, Vac Filt.	Thickening, Anaer. Dig.	Thickening, Anaer. Dig	Filter Press, Incineration	DAF Thickening	Anaer Dig.
Strippers, no.	1	1	1	2	4	2	5
Reactor-Clarifiers or <u>M</u> ixer/ <u>F</u> locculators, no	MF 1	RC 1	MF 1	RC 2	RC 4	RC 1	MF 2
Elutriation source ²	SR	RC/SEC	PR	RC	RC	RC	SR

¹ With final final filtration; chemical polishing available but not utilized

² Sludge Recycle elutriation; Reactor-Clarifier overflow elutriation, PRImary effluent supplement; SECONDary effluent.

Table 3-5. Performance Data Summary for Full-scale Phostrip Plants.

Plant	Design Flow	Startup Date	Data Period	Total Phosphorus, mg/l						Notes	Ref.
				Influent Averages			Effluent Averages				
				min. mo.	ave. mo.	max. mo.	min. mo.	ave. mo.	max. mo.		
	m ³ /d		mo.								
Seneca Falls, NY	3,400	1973	1		6.3			0.6		Full-scale demo	6
Landsdale, PA	9,500	1982	12	4.0	5.2	6.4	0.6	1.2	2.0		58
Adrian, MI	26,500	1981	11	3.4	4.4	5.3	<0.1	0.4	0.6	Excludes periods of upset due to other plant problems	59
Savage, MD	56,800	1982	6	5.7	8.1	9.3	0.5	1.2	1.7		58
			1		6.6			1.7		July 1984	47
			1		7.0			0.5		April 1985	47
Southtowns, NY	60,600	1982	4	2.3	3.2	4.1	0.3	0.5	0.9		59
Amherst, NY	90,900	1982	12	2.9	5.2	14.3	0.4	1.3	2.5		59
Reno, NV	113,700	1981	4		7.0-7.3			0.8-1.1		9/82-12/82	60

precipitated solids from a sludge lagoon (58). Advanced control instrumentation and problems with the lime feeder also added to startup difficulties. The method of introducing the recycle sludge to the stripper tank inlet was modified to prevent the entrainment of air that would counteract anaerobic conditions in the stripper. A distributor was devised to introduce the return sludge at a depth of 1.8 m (6 ft.).

3.3.2 Modified Bardenpho Process Performance

Table 3-6 summarizes basic design parameters for the first two full-scale North American Modified Bardenpho facilities operating in the United States at Palmetto, Florida, and in Canada at Kelowna, British

Columbia, respectively. Both plants are operated with relatively long detention times to provide sufficient volume for nitrification and denitrification as well as biological phosphorus removal. The design SRT for the Kelowna plant is 1.5-2.0 times the Palmetto plant due to the need to operate at much colder wastewater temperatures. Both plants have polishing filters to meet stringent effluent requirements for BOD and suspended solids as well as nutrients. Required effluent limits for the Palmetto plant are equal to or less than 5, 5, 3, and 1 mg/l for TBOD, total suspended solids (TSS), total nitrogen, and total phosphorus, respectively. The allowable limits for TBOD, TSS, total nitrogen, and total phosphorus for

Table 3-6. Modified Bardenpho Process Full-Scale Plant Design Summary.

	Palmetto, FL (64)	Kelowna, Canada (62,63)
Startup date	10/79	5/82
Flow, m ³ /d	5,300	22,700
Detention time, hr (no. cells)		
Anaerobic zone	1.0 (1)	2.0 (1)
Anoxic 1 zone	2.7 (1)	4.0 (4)
Nitrification zone	4.7 (1)	9.0 (9)
Anoxic 2 zone	2.2 (1)	4.0 (4)
Reaeration zone	1.1 (1)	2.0 (2)
Total, hr	11.6	21.0
SRT, days	20	30-40
MLSS, mg/l	3,500	3,000
Temperature, °C	18-25	9-20
Sec. clarifier application rate, m ³ /m ² /d	22.3	14.0
Polishing filter application rate, m ³ /m ² /d	93.7	23.4
Primary treatment	No	Yes
Biological sludge handling	Drying Beds	DAF Thick., Composting

the Kelowna plant are 8, 7, 6, and 2 mg/l, respectively.

By coincidence, both plants use submerged turbine devices for aeration. Other Modified Bardenpho plants are using different aeration devices such as fine bubble diffusion at Payson, Arizona, and at some South African plants (57,65). The Carrousel oxidation ditch system has been used in the Modified Bardenpho designs at Fort Meyers and Orange County, Florida. The Kelowna plant was designed with submerged turbine mixers and/or aerators in each of the multiple cells used in the various anaerobic, anoxic or aerobic zones to provide maximum operating flexibility. The nitrification zone, for example, has nine cells with each stage having a design detention time of 1 hour. The last four cells can be operated with or without air addition with the turbine mixers, so that the nitrification detention times and second anoxic zone detention times can be varied. The second anoxic zone also can be operated with air addition to provide additional nitrification time, if needed.

Both plants shown in Table 3-6 use low-head, automatic backwash filtration to meet effluent suspended solids requirements. Both plants were designed with primary clarification, but the Palmetto plant modified its operation after startup (16) by introducing the return activated sludge to the primary tank to effectively use the tank to provide additional anaerobic fermentation time. At the Kelowna plant some of the organic material in the solids removed in primary treatment can also be used to support the

biological phosphorus removal process. The supernatant from the primary sludge gravity thickener can be returned to the anaerobic fermentation zone or first anoxic zone. This supernatant contains fermentation products when the solids are held in the thickener long enough.

Another similarity with the two plants is that they do not employ either aerobic or anaerobic digestion of the waste activated sludge. The practice of sludge digestion has not been recommended for Modified Bardenpho biological phosphorus removal plants (12) since the recycle of phosphorus expected to be released during sludge destruction would overload the phosphorus removal capacity of the activated sludge system. At Palmetto, the sludge is wasted from the sludge recycle line directly to drying beds. The recycle flow from the drying bed underdrains results in a minimal additional phosphorus load to the plant (66). The ultimate disposal of the drying bed sludge is by land application. At Kelowna, dissolved air flotation thickening of the waste activated sludge is used to prevent anaerobic conditions and phosphorus release. The sludge is further composted prior to ultimate disposal.

One year of plant performance data (April 1981-March 1982) is shown for the Palmetto facility in Table 3-7. The data are monthly average summaries of influent wastewater and filtered effluent quality obtained from plant records. Due to the relatively weak influent wastewater, biological phosphorus removal capacity was limited. As shown, alum was added prior to the secondary clarifier during 5 months of the 1-year period to reduce the orthophosphorus concentration and to achieve an effluent total phosphorus concentration generally less than 1 mg/l. The plant also consistently achieved effluent total nitrogen concentrations of less than 3 mg/l.

Table 3-8 shows a 2-year performance summary for the Kelowna plant from January 1983 through December 1984. During this period, a number of operating conditions were investigated and the influent load was at about 54 percent of the design load. The final filter effluent summary for the 2-year operation includes results for a two-train operation, a single-train operation, thickener supernatant feeding, and splitting of the return sludge between the fermentation zone and first anoxic zone. The plant effluent was well within the treatment requirements for nitrogen and phosphorus removal. During the one-train operation period (May - November, 1984), the single train was loaded at about 120 percent of its design loading and effluent requirements were met.

3.3.3 A/O Process Performance

The first full-scale demonstration of the A/O biological phosphorus removal process occurred with the conversion of one-third of the 34,000-m³/d (9-mgd) Largo, Florida contact stabilization system

Table 3-7. Palmetto, FL Modified Bardenpho Process Performance (April 1981 - March 1982 Monthly Averages) (49).

	April	May	June	July	Aug.	Sept.	Oct	Nov.	Dec.	Jan.	Feb.	March
<u>Influent</u>												
Flow, m ³ /d	3,200	3,000	3,500	3,600	5,500	5,900	3,700	3,300	3,200	3,600	3,700	4,500
TBOD, mg/l	164	159	124	104	74	67	113	157	182	160	163	150
TSS, mg/l	155	157	144	112	76	76	116	160	182	141	167	128
Temperature, °C	25	27	29.5	30.5	29.5	29	28	27	24	23	23	23
TKN, mg/l	31.8	40.8	30.1	25.0	19.7	21.9	28.1	40.0	38.2	37.7	42.4	32.4
NH ₄ -N, mg/l	25.0	25.2	20.4	18.7	12.7	12.7	17.8	22.6	27.2	28.0	25.8	23.7
Total P, mg/l	9.2	6.4	7.0	5.6	4.1	4.9	6.3	8.5	8.8	8.7	8.0	6.6
Ortho P, mg/l	6.5	6.1	5.3	4.5	2.8	3.5	4.7	5.9	5.9	5.4	5.2	4.4
Alkalinity, mg/l	174	169	156	154	143	140	144	171	198	191	201	187
<u>Filtered effluent</u>												
TBOD	2	1	1	1	1	1	1	1	1	1	1	1
TSS, mg/l	3	2	2	2	2	1	2	2	2	1	2	3
Total N, mg/l	2.1	2.1	1.9	2.8	2.0	1.7	1.9	2.1	2.5	2.7	2.6	2.8
NO ₃ -N, mg/l	1.0	1.3	1.0	1.9	1.2	1.1	1.1	1.3	1.5	1.9	1.8	2.1
NH ₄ -N, mg/l	0.4	0.3	0.3	0.2	0.2	0.2	0.2	0.2	0.4	0.3	0.1	0.2
Total P, mg/l	2.5	3.4	2.6	1.8	1.5	1.2	1.1	0.7	1.6	0.6	0.8	0.9
Ortho P, mg/l	2.2	1.4	2.5	1.7	1.1	1.3	0.9	0.7	1.0	0.5	0.7	0.8
							*	*		*	*	*

* Minimal alum dose applied prior to secondary clarification.

Table 3-8. Kelowna, Canada Modified Bardenpho Process Performance Results of Cumulative Frequency Plot of Data (62).

	Median	Lower 5%	Upper 5%
<u>Influent</u> (1/83 - 12/84)			
Flow, m ³ /d	12,400	10,400	10,000
COD, mg/l	3.3	2.7	2.6
TKN, mg/l	24.5	19.0	33.5
NH ₄ -N, mg/l	17.5	15.0	21.1
Total P, mg/l	4.5	3.3	5.8
Ortho P, mg/l	3.8	3.0	4.3
<u>Final Effluent</u> (1/83 - 12/84)			
TKN, mg/l	1.5	0.2	1.8
NO ₃ -N, mg/l	1.8	0.8	4.2
NH ₄ -N, mg/l	0.1	<0.1	6.0
Total P, mg/l	0.8	0.2	1.8
Ortho P, mg/l	0.77	0.15	2.25
<u>Final Effluent - 1 Train</u> (5/22/84-11/9/84)			
Flow, m ³ /d	14,000	12,000	17,000
NO ₃ -N, mg/l	2.0	1.2	3.4
NH ₄ -N, mg/l	<0.1	<0.1	0.75
Ortho P, mg/l	1.1	0.08	1.75

to an A/O system treating primary effluent. The total average hydraulic detention time was 4.1 hours with 0.9 hour in three anaerobic stages, 0.6 hour in two anoxic stages, and 2.6 hours in five aerobic stages (29). The A/O system had its own secondary clarifier. During nitrification operating conditions, mixed liquor from the fifth stage of the aerobic zone was recycled back to the first anoxic stage. The MLSS concentration and SRT were decreased to provide a non-nitrifying operation. At that condition, the two anoxic stages were operated as additional anaerobic stages and internal recycle of mixed liquor was not practiced.

Table 3-9 summarizes effluent concentrations reported for nitrifying and non-nitrifying conditions. Slightly lower TBOD and higher TSS concentrations are shown for the nitrification operating period. The orthophosphorus and total phosphorus effluent concentrations were also higher. Unfortunately, insufficient data were presented to allow a determination of whether the difference in performance was related to the nitrification operation or changes in the primary effluent fed to the A/O system. Data on influent characteristics for another time period (August 1979-July 1980) indicated that the TBOD, total phosphorus, and TKN ranges were 81-127 mg/l, 7.0-9.4 mg/l, and 21.4-30.5 mg/l, respectively (29). The influent SBOD was about 50 percent of the influent TBOD.

Table 3-9. Summary of A/O Process Effluent Quality (Average Monthly and Monthly Range), Largo, FL (29).

Parameter	Non-Nitrification	Nitrification
Data period	2/81 - 6/81	9/81 - 2/82
TBOD, mg/l	7 (6-8)	5 (4-7)
TSS, mg/l	10 (8-13)	18 (10-22)
Total P, mg/l	1.4 (1.2-1.5)	1.7 (1.3-2.2)
Ortho P, mg/l	0.6 (0.5-0.8)	1.0 (0.5-2.0)

The phosphorus content measured for the waste activated sludge was reported to be 4.2-6.0 percent (17). The sludge handling method proposed for the Largo facility consisted of dewatering, drying, and pelletizing for use as a soil conditioner/fertilizer.

Table 3-10 describes operating conditions for a two-train A/O process at a Pontiac, Michigan EPA demonstration site. The original activated sludge system consisted of four plug flow trains with coarse bubble diffusion. Two of the trains were converted to the A/O process to allow a performance comparison of biological phosphorus removal to conventional activated sludge treatment. The existing tankage was divided into desired stages by the installation of wooden baffles. Side-mounted submersible mixers were installed to provide mixing in the anaerobic stages after plugging the diffuser lines. The demonstration project also allowed an evaluation of the A/O process under cold temperature operation, during nitrification, and with anaerobic digestion supernatant return (67).

Table 3-10. Pontiac, MI A/O System Operating Conditions (67).

Operating dates	7/13/84 - 3/31/85
Average flow, m ³ /d	12,200
Average detention time, hr (no. cells)	
Anaerobic zone	1.8 (3)
Aerobic zone	6.7 (4)
SRT, days	16-24
Temperature, °C	10-17
Primary treatment	Yes
Biological sludge handling	Anaerobic Digestion

The treatment performance for different operating phases is summarized in Table 3-11. Effluent total phosphorus concentrations averaged less than 1.0 mg/l during the study period, with complete or partial nitrification. Influent phosphorus concentrations were relatively low, and a comparison of performance with a parallel conventional activated sludge system showed an increased total phosphorus removal of 1.2-1.6 mg/l for the A/O process operation. Other factors that could have potentially limited biological

phosphorus removal performance at Pontiac were the relatively long SRT for an A/O system and the recycle of nitrate nitrogen in the return sludge to the anaerobic zone. It appears there was sufficient BOD in the influent to offset these considerations. Evaluation of the digester supernatant during this study indicated a minimal level of soluble phosphorus was released during anaerobic digestion. Consequently, the impact of supernatant recycle on system performance was minimal. One explanation offered for this was the possibility of the formation of a magnesium ammonium phosphate precipitate, but further study was suggested. Average effluent TSS concentrations from the secondary clarifier ranged from 6 to 10 mg/l, thereby minimizing the concentration of particulate phosphorus in the final effluent.

Table 3-11. Full-scale A/O Process Performance, Pontiac, MI (67).

Influent	Phase I	Phase II*	Phase III	Phase IV*
<u>Influent</u>				
Flow, m ³ /d	11,300	10,800	12,070	14,680
TBOD, mg/l	110	137	143	112
SBOD, mg/l	65	65	87	65
NH ₄ -N, mg/l	15.2	17.8	16.1	18.5
Total P, mg/l	3.2	4.1	3.7	3.0
Soluble P, mg/l	1.9	2.2	2.2	1.6
Temperature, °C	17	16	11	10
<u>Reactor</u>				
MLSS, mg/l	2,820	2,410	2,340	2,360
MLVSS, mg/l	1,800	1,670	1,640	1,590
SRT, days	24	21	19	16
<u>Effluent</u>				
TBOD, mg/l	6.2	9.4	12.9	12.7
SBOD, mg/l	1.8	3.0	2.6	2.0
NH ₄ -N, mg/l	0.9	2.8	5.9	4.5
NO ₃ -N, mg/l	10.4	11.6	6.7	8.8
Total P, mg/l	0.8	0.7	0.4	0.7
Soluble P, mg/l	0.7	0.6	0.3	0.5
TSS, mg/l	6	7	8	10
VSS, mg/l	4	4	5	6

* Anaerobic digester supernatant returned during these phases.

3.3.4 Operationally Modified Activated Sludge Process Performance

An operationally modified activated sludge system involves turning off aerators at the front end of the activated sludge basin to create anaerobic fermentation conditions to provide the preferred substrate for the phosphorus-removing bacteria. The early reports on "luxury uptake" of phosphorus were for plug flow plants at Baltimore, Maryland; San Antonio, Texas; and Los Angeles, California. Oxygen transfer with the diffused aeration systems at the front end of these activated sludge systems was apparently sufficiently limited to stimulate anaerobic fermentation conditions. Unaerated conditions were purposefully

created at the front end of activated sludge aeration basins at DePere, Wisconsin, and at the Reedy Creek Improvement District main plant in Lake Buena Vista, Florida. The performance of these operationally modified plants has been evaluated and the results reported (47).

The Reedy Creek plant serves the Walt Disney World resort complex. Operating conditions for the plant are shown in Table 3-12. The plant uses four parallel plug flow aeration basins following primary treatment. The initial third of each basin is unaerated. Backmixing provides sufficient agitation to maintain suspension of solids in the unaerated zone. Nitrification and denitrification also occur in the system.

Table 3-12. Operating Conditions for Operationally Modified Activated Sludge Systems (47).

	Reedy Creek	DePere
Design flow, m ³ /d	22,700	53,750
Detention time, hr		
Unaerated zone	3.0	7.5
Aerated zone	6.0	15.0
Sec. clar. overflow rate, m ³ /m ² /d	14.7	17.9
SRT, days	7.2	10.6
MLSS, mg/l	2,100	3,000
Return sludge ratio	0.59	0.81
Primary treatment	Yes	No
Sludge handling	DAF Thick., Aerobic Dig., Land Spread	DAF Thick., Filter Press, Incineration

The DePere operation involved modifying a contact-stabilization activated sludge system. The stabilization tank air supply was stopped, and mixing was accomplished in the basin with turbine aerator mixers. A complete mix aerated contact basin was then used for the aerobic treatment step. The detention times are given in Table 3-12. The plant was operating at about 50 percent of its design capacity, resulting in the relatively long detention times shown.

Table 3-13 indicates that both plants achieved relatively low effluent phosphorus concentrations during the three-month summer test period. Effluent suspended solids concentrations were low enough to minimize the contribution of particulate phosphorus to the final effluent. Nitrification was occurring in both plants, but apparently the biological phosphorus removal levels were not limited by the nitrate nitrogen present in the return activated sludge. This may have been due to the level of BOD in the influent, warm wastewater temperatures, and relatively long unaerated detention times.

Table 3-13. Average Performance of Operationally Modified Activated Sludge Systems (47).

	Reedy Creek	DePere
Test dates: June - August 1984		
<u>Influent</u>		
TBOD, mg/l	155	150
SBOD, mg/l	85	86
Total P, mg/l	6.7	5.1
Ortho P, mg/l	5.3	1.9
<u>Effluent</u>		
TBOD, mg/l	3	7
TSS, mg/l	13	7
Total P, mg/l	0.9	0.3
Ortho P, mg/l	0.4	0.1
NH ₄ -N, mg/l	0.7	1.4

3.3.5 Factors Affecting Performance

3.3.5.1 Phostrip Process Solids Detention Time and Elutriant Quality

In evaluating effects on performance, one must distinguish between the sidestream operation of the Phostrip system and the mainstream systems. The Phostrip system has shown the highest degree of treatment flexibility and treatment effectiveness with low organic strength wastewaters because a substantial amount of the total phosphorus removal can occur via the stripper and chemical precipitation operations. For example, effluent total phosphorus concentrations of less than 1 mg/l were obtained at Lansdale, Maryland, in spite of an average influent TBOD of only 41 mg/l and an influent total TBOD:total phosphorus (TP) ratio of only about 8.1. Under such conditions the mainstream biological phosphorus removal processes would be expected to achieve less efficient phosphorus removal.

The critical design and operating parameters that affect performance in the Phostrip process are the stripper SDT, the elutriation rate, and the elutriant source. The proposed mechanism for biological phosphorus removal suggests that sufficient SDT is needed in the stripper to form substrate fermentation products from lysed bacteria. Successful performance has been observed for stripper SDT values in the range shown in Table 3-2. Longer SDT values are suggested for operations with a significant quantity of oxidized nitrogen entering the stripper, either via the return activated sludge stream or the elutriant stream. For such cases, a 50-percent increase in the stripper SDT has been recommended (48).

The elutriant source can affect the SDT design of the stripper and overall performance. The least desirable elutriant source would be a nitrified secondary effluent with a significant DO level. Some of the available substrate in the stripper operation would be needed to reduce the DO and nitrate oxygen before the

necessary organic fermentation activity could occur. As conditions change in the plant, the SDT may be adjusted in the stripper by varying the sludge blanket depth. The chemical treatment system overflow has been frequently used as a stripper elutriant source because of its low phosphorus content. Another potential elutriant source is primary effluent. The readily available organic material in primary effluent could result in a lower SDT, since less organic material is needed from the lysing of biological solids in the return activated sludge. This elutriant source should contain little, if any, DO and no oxidized nitrogen.

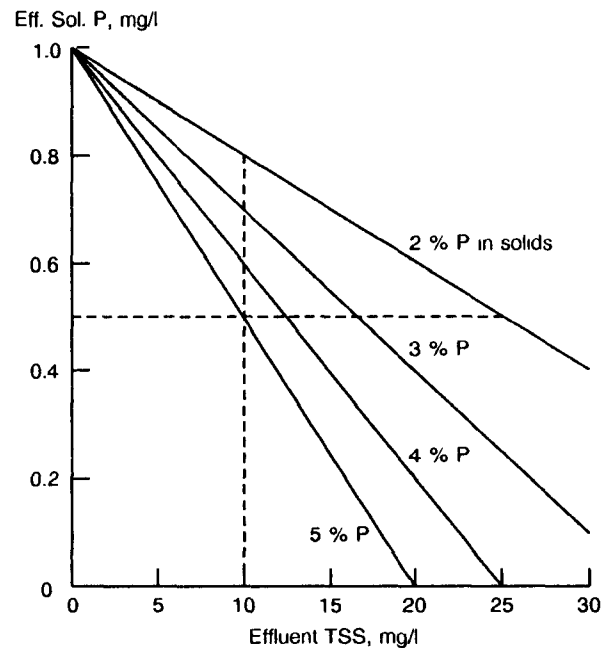
3.3.5.2 Effluent Suspended Solids

An important plant performance consideration for both the Phostrip and mainstream processes is the secondary effluent suspended solids concentration and phosphorus content of those solids. This is even more critical for the mainstream processes because they normally produce mixed liquor suspended solids higher in phosphorus content than the Phostrip process. Phosphorus contents of MLSS on a dry solids basis of 2.3-5.8 percent have been reported for Phostrip and mainstream systems (47), with values in the lower portion of this range reported for the Phostrip process.

Many of the nutrient removal facilities, including Palmetto, Kelowna, and Payson, have required final filtration to meet very low effluent suspended solids and BOD limits as well as nutrient removal requirements. In other cases, low effluent suspended solids may not be required. In these cases, the necessity of effluent filtration may be evaluated as a means to meet the required effluent total phosphorus concentration. This consideration is illustrated in Figure 3-9. An effluent total phosphorus concentration requirement of 1.0 mg/l is assumed. If the effluent soluble phosphorus concentration is 0.5 mg/l and the phosphorus content of the MLSS is 5 percent, the effluent TSS concentration has to be 10 mg/l or less to meet the 1.0-mg/l effluent total phosphorus limit. If the solids phosphorus content were 3 or 4 percent, the effluent TSS would have to be equal to or less than 17 or 12.5 mg/l, respectively. Thus, unless excellent secondary clarifier performance is achieved or the effluent soluble phosphorus concentration is very low (e.g., 0.2 mg/l or less), a polishing filter would be required to meet an effluent total phosphorus level of 1.0 mg/l. With the exception of the Largo nitrification operating condition, the full-scale plants described in the performance section were able to produce effluent TSS concentrations of less than 15 mg/l. The effluent suspended solids data shown for Palmetto were after filtration, but effluent TSS concentrations of 4-8 mg/l from the secondary clarifier have been reported (16). It appears that a conservatively designed secondary clarifier could produce effluent suspended solids concentrations low enough to meet a typical effluent

total phosphorus requirement of 1.0 mg/l, provided the soluble phosphorus concentration in the effluent stream does not exceed 0.4-0.6 mg/l.

Figure 3-9. Maximum effluent soluble P concentration for effluent total P < 1.0 mg/l.



3.3.5.3 Available Organics for Phosphorus Removal

Effluent soluble phosphorus concentrations as low as 0.1-0.2 mg/l have been achieved in the Modified Bardenpho, A/O, and operationally modified activated sludge processes (47,62,68). However, this is not achieved at all plants and for all operating conditions because of the dependence on the availability of fermentation substrate products needed by the phosphorus-storing bacteria relative to the amount of phosphorus that must be removed in the system. In addition, as will be discussed further, the required ratio of fermentation substrate per unit of phosphorus removed is affected by the amount of nitrate nitrogen entering the fermentation zone and also by the SRT of the system.

Based on the biological phosphorus removal mechanism described in Section 3.1, a given amount of fermentation products, such as acetate, consumed by the phosphorus-removing organisms will yield a certain quantity of new organisms. A significant fraction of the dry weight of these organisms will be phosphorus, and phosphorus removal eventually occurs in the system by the wasting of these organisms. Fundamental studies working with pure cultures of *Acinetobacter* have determined a synthesis yield of 0.42 g solids/g acetate and a phosphorus content of 6-10 percent (69,70). Thus, if

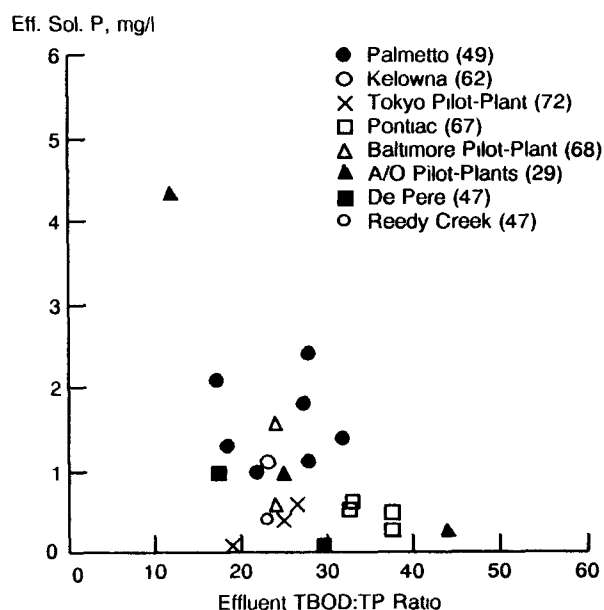
the amount of acetate or similar fermentation products that could be consumed in a biological phosphorus system were known, the quantity of phosphorus that could be removed could be estimated. Unfortunately, the complexity of the process has thus far prevented the determination of the amount of fermentation products produced, and then consumed, by the phosphorus-storing organisms. The fermentation products used by the phosphorus-storing organisms will be generated in the anaerobic zone and some may be present in the influent of more septic wastewaters. Due to the rapid assimilation of the fermentation products in the anaerobic zone, it has not been possible to measure their production rate (70).

Since the amount of fermentation products produced in the system can not be measured, other indirect methods have been proposed in an attempt to quantify the phosphorus removal potential of a system. Siebritz *et al.* (71) recognized that municipal wastewater is made up of slowly biodegradable substrate and a substrate fraction that is biodegraded more rapidly. They measured the immediate oxygen uptake of a mixed liquor upon addition of a wastewater sample to quantify what they termed the readily degradable portion of the substrate. They established a minimum influent concentration of readily degradable substrate of 25 mg/l for biological phosphorus removal to proceed. Nicholls *et al.* (70) had difficulty relating such measurements to biological phosphorus removal performance. They proposed an alternate method using nitrate as the electron acceptor. In this method, the initial zero-order nitrate reduction rate after addition of a wastewater sample to the mixed liquor is measured. The amount of nitrate nitrogen used during the time that a zero-order reduction occurs is then related to a readily biodegradable substrate by a stoichiometric conversion. The basic concept in both methods is that the more readily biodegradable substrate may be the source of the fermentation products in the anaerobic zone of biological phosphorus removal systems.

As an alternate to these special tests, Hong *et al.* (29) have used the soluble BOD concentration of the influent wastewater as an indication of the amount of substrate readily available for the formation of fermentation products. They have recommended an influent SBOD: soluble phosphorus (SP) ratio of at least 15 to produce an effluent soluble phosphorus concentration below 1.0 mg/l for A/O systems operating at F/M loadings above 0.15 kg TBOD/kg MLVSS/d. Data presented by Tetreault *et al.* (47) from the full-scale Largo A/O system operation supported this recommendation. At influent SBOD:SP ratios below 12, effluent soluble phosphorus concentrations varied from 0.5 to 4.5 mg/l.

Influent SBOD has not been measured at many of the full-scale biological phosphorus removal plants for a variety of possible reasons. Such reasons include the facts that the mechanism of biological phosphorus removal has only recently begun to be unraveled and the relative degrees of soluble and particulate BOD fermentation are not known. It has been recognized that more phosphorus removals and lower effluent soluble phosphorus concentrations occur for wastewaters with higher influent TBOD:TP ratios. Figure 3-10 summarizes data showing effluent soluble phosphorus concentrations and TBOD:TP ratios. Tetreault *et al.* (47) have recommended a TBOD:TP ratio of greater than 20-25 to achieve an effluent soluble phosphorus concentration below 1.0 mg/l.

Figure 3-10. Effluent soluble P concentration vs. influent TBOD:TP ratio.



3.3.5.4 Effect of Solids Retention Time

Figure 3-10 shows that higher effluent soluble phosphorus concentrations occurred for the Modified Bardenpho facilities operated within the same range of influent TBOD:TP ratios as other biological phosphorus removal designs. The Bardenpho systems were operated, as expected, at longer SRTs to accomplish nitrification and denitrification. Lower sludge yields associated with the longer SRTs would logically decrease the phosphorus removal capacity for the system.

Assuming a 4.5-percent waste activated sludge phosphorus content, Barth and Stensel (66) suggested a TBOD removal:TP removal ratio of 33 at an SRT of 25 days and a ratio of 25 at an SRT of 8 days. Fukase *et al.* (72) found, in an A/O system

pilot-plant study treating municipal wastewater, that the TBOD removal:TP removal ratio increased from 19 to 26 as SRT was increased from 4.3 to 8.0 days. At the same time, the phosphorus content of the activated sludge decreased from 5.4 to 3.7 percent.

Maier *et al.* (73) found in pilot-plant studies that the rate of phosphorus uptake per unit of mixed liquor solids decreased by a factor of 2.6 as the F/M loading was decreased from 0.2 to 0.1 kg TBOD/kg MLVSS/d. Tracy and Flammiano (74) showed that for identical influent TBOD:TP ratios of 16, the rate of phosphorus uptake in the aerobic zone decreased by a factor of 3 as the F/M loading was decreased from 0.44 to 0.24 TBOD/kg MLVSS/d in bench-scale studies.

These results indicate that operation at longer SRT values will decrease the efficiency of phosphorus removal per unit of BOD removed. To maximize biological phosphorus removal, systems should not be operated with SRT values in excess of that required for overall treatment needs. Systems that require nitrification and denitrification, such as the Modified Bardenpho system or extended aeration systems promoting sludge stabilization, will require much higher influent TBOD:TP ratios to produce soluble phosphorus concentrations below 1.0 mg/l.

3.3.5.5 Nitrate Nitrogen in the Anaerobic Zone

Barnard (12) was the first to point out that nitrate nitrogen entering the anaerobic zone of biological phosphorus removal systems could reduce the phosphorus removal capability of the system. He attributed this to an increase in the redox potential of the reactor and a reduction in the degree of anaerobic stress to induce phosphorus release. However, an improved understanding of the phosphorus removal mechanism indicates that nitrate reduction in the anaerobic zone utilizes substrate that would otherwise be available for assimilation by the phosphorus-storing organisms. Thus, nitrate has the effect of reducing the net influent BOD:P ratio for the system. Because of this, variable results have been observed for systems with nitrate nitrogen present. The degree of variability depends on the system influent BOD and phosphorus concentrations and the system SRT. The return activated sludge recirculation ratio is also important as this affects the amount of nitrate nitrogen fed to the anaerobic zone.

Simpkins and McClaren (55) reported a total phosphorus removal efficiency reduction from 90 to 55 percent when the effluent nitrate nitrogen concentration increased from 4.0 mg/l to 6.7 mg/l in a Modified Bardenpho pilot-plant study. During the Palmetto Modified Bardenpho operation (16), the internal recycle pumps were stopped, causing the effluent nitrate nitrogen concentration to increase to about 10 mg/l and the effluent total phosphorus concentration to increase from 2.3 to 7.1 mg/l. These

two examples illustrate the sensitivity of phosphorus removal efficiency to nitrate nitrogen entering the anaerobic zone of long-SRT biological phosphorus removal systems treating relatively weak wastewaters.

Similar effects of nitrate nitrogen on biological phosphorus removal have been reported in an operationally modified activated sludge system (13). Vinconneau *et al.* (75) also showed that nitrate could significantly affect biological phosphorus removal performance for a lightly loaded A/O system. At similar influent BOD:P ratios and operating F/M loadings, the effluent total phosphorus concentration decreased from 2.0 to 0.9 mg/l as the effluent nitrate nitrogen concentration decreased from 3.4 to 0.6 mg/l. The effluent soluble phosphorus concentration at Pontiac (67) was consistently below 1.0 mg/l even though the average effluent nitrate nitrogen concentration ranged from 6.7 to 11.6 mg/l for the four study periods reported. The low effluent phosphorus concentration was attributed to a relatively high influent BOD:P ratio so that excess BOD was available to reduce the nitrate. The operationally modified activated sludge systems at Reedy Creek and De Pere (47) also had a relatively high influent BOD:P ratio and produced low effluent soluble phosphorus concentrations in spite of the occurrence of nitrification. In the Reedy Creek system, nitrate nitrogen was also fed to the anaerobic zone by internal circulation of the nitrified mixed liquor as well as via the return activated sludge. Nitrification and denitrification were occurring concurrently as indicated by effluent nitrate nitrogen concentrations of less than 4 mg/l and ammonium nitrogen concentrations of less than 1 mg/l.

Rabinowitz (34) studied the effect of nitrate nitrogen concentration on phosphorus release in batch tests using activated sludge developed in a UCT system pilot plant. Sodium acetate was used for the substrate source. He found that with excess substrate available, the phosphorus release during anaerobic contacting was inversely proportional to the amount of nitrate nitrogen present. He further found that the denitrification of nitrate in the anaerobic batch tests had the effect of reducing the availability of substrate for phosphorus release. The substrate consumption for denitrification was found to be 3.6 mg COD/mg nitrate nitrogen reduced. This ratio is in close agreement with a ratio of 3.53 developed by McCarty (76) for denitrification using acetate. The ratio in actual wastewater treatment systems will depend on the characteristics of the substrate used for denitrification in the anaerobic zone. With nitrate nitrogen present, substrate would not necessarily be converted to volatile fatty acids by fermentation but could be used for denitrification directly. A substrate consumption ratio determined from an anoxic-aerobic pilot-plant system treating domestic wastewater was about 5.0 mg soluble COD/ mg nitrate nitrogen reduced for complete denitrification

(49). The same reference reported on substrate consumption for anoxic-aerobic system denitrification using eleven different industrial wastewater substrate sources. The mean substrate consumption ratio was 5.3 mg COD/mg nitrate nitrogen reduced, with reported values of 2.2-10.2.

Some investigators have also reported on experiments that suggested biological phosphorus-removing organisms are capable of denitrification. Under substrate limiting conditions, phosphorus uptake and nitrate reduction occurred simultaneously in an anoxic reactor (15,34). Nitrate reduction was also observed in a pure culture experiment with *Acinetobacter* 210A (69). The nitrate nitrogen was reduced only to nitrite nitrogen, which was toxic at about a 500 mg/l concentration.

3.3.5.6 Wastewater Temperature

The Phostrip, Modified Bardenpho, and A/O processes have been applied successfully for both cold and warm wastewater temperature conditions. Reported data on the operation and performance of operationally modified activated sludge systems for biological phosphorus removal during cold wastewater temperature conditions are limited.

Peirano *et al.* (77) reported that wastewater temperature had no significant effect on Phostrip process efficiency during plant-scale testing at Reno-Sparks. This is likely the result of having an adequate size stripper to handle lower activity levels at cold temperatures. Shapiro *et al.* (4) showed specific phosphorus release rates for activated sludge ranging from 0.63 mg/l-hr/g of volatile suspended solids at 10°C (50°F) to 3.15 at 30°C (86°F).

Modified Bardenpho systems have been designed at about twice the total hydraulic detention time for treatment of 10°C (50°F) wastewater vs. 20°C (68°F) wastewater (63,64). This difference is due to the effect of temperature on the nitrification-denitrification design and is not related to the phosphorus removal design. Prior to the Kelowna plant design, bench-scale studies showed that 90 percent biological phosphorus removal was possible over a temperature range from 18°C (64°F) down to 6°C (43°F) (78). The study did exhibit a decreased nitrogen removal efficiency below 10°C (50°F), however.

The full-scale A/O system operation demonstrated at Pontiac, Michigan, revealed that biological phosphorus removal was not affected by wastewater temperatures as low as 10°C (50°F) (67). Biological phosphorus removal was studied in laboratory batch units over a temperature range of 5-15°C (41-60°F) by Sell *et al.* (46). The amount of phosphorus removed at 5°C (41°F) vs. 15°C (60°F) was greater by more than 40 percent. The improvement was credited to a population shift to more slow growing

psychrophilic bacteria with a higher cell yield. Groenestijn and Deinema (69) reported that the phosphorus content of a pure culture of *Acinetobacter* decreased from 10.1 percent at 5°C (41°F) to 1.4 percent at 35°C (95°F). An A/O system operating at a low organic loading rate of 0.032 kg COD/kg MLSS/d produced its lowest effluent soluble phosphorus concentrations of 0.9 mg/l during the coldest operating month when wastewater temperature was 5°C (41°F). The phosphorus content of the sludge was 4.7 percent compared to a range of 3.5 to 4.9 percent for five other months (75).

3.3.5.7 pH

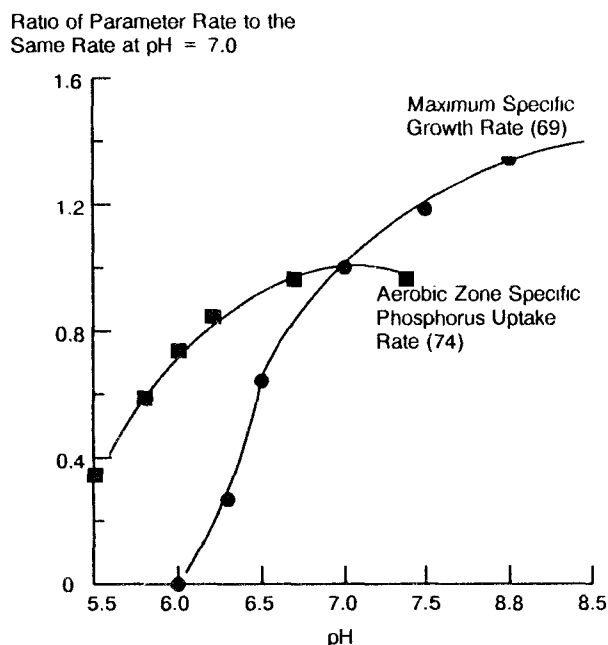
Two different laboratory experiments using synthetic wastewater to evaluate the effect of pH have been reported. Groenestijn and Deinema (69) studied the effect of pH at a wastewater temperature of 25°C (77°F) on the maximum specific growth rate of a strain of *Acinetobacter*. As shown in Figure 3-11, the maximum specific growth rate was 42 percent higher at a pH of 8.5 compared to that at a pH of 7.0. Below a pH of 7.0, a steady decline in the maximum specific growth rate occurred. Below a pH of 6.0 the organisms did not grow. Between pH values of 6.5-8.0, the phosphorus content of the culture remained constant at about 6.0 percent. It increased to 7.5 percent at a pH of 6.0. Tracy and Flammio (74) studied the effect of pH on the specific phosphorus uptake rate (g P/g VSS/hr) in the aerobic phase of an anoxic-aerobic lab reactor. Their results are also normalized to a pH of 7.0 in Figure 3-11. They claimed little difference in the phosphorus uptake rate from a pH of 6.5 to a pH of 7.0. Below a pH of 6.5, the phosphorus uptake rate declined steadily. They further stated that all activity was lost at a pH of 5.2. As they increased the pH, they claimed that the phosphorus uptake activity was essentially duplicated.

Nagashima *et al.* (79) found that total phosphorus removal in the Modified Bardenpho process was improved from 42 to 92 percent as the pH was increased from 5 to 8. These results suggest that the efficiency of biological phosphorus removal may decline significantly below a pH of 6.5.

3.3.5.8 DO Concentration in the Phosphorus Uptake Zone

No specific studies have been reported that address the effect of the DO concentration on biological phosphorus removal. The biological phosphorus removal mechanism suggests that the DO concentration may affect the rate of phosphorus uptake in the aerobic zone, but not the amount of phosphorus removal possible, provided that sufficient aerobic time is available. The mechanism teaches that the oxidation of stored or exogenous carbonaceous materials produces energy for the incorporation of soluble phosphorus into cellular polyphosphate compounds. In the treatment of an acetate wastewater in an anaerobic-aerobic fill-

Figure 3-11. Reported effects of pH on biological phosphorus removal.



and-draw system, Fukase *et al.* (32) showed that aerobic detention times required for maximum phosphorus uptake were 1-2 hours. Tracy and Flammino (74) showed that 80 minutes was required for the A/O process treating a municipal wastewater and 160 minutes for treating a municipal-food processing wastewater combination.

Miyamoto-Mills *et al.* (38) obtained effluent total phosphorus concentrations below 1 mg/l in a Phostrip system pilot-plant study with the aerobic stage operating at DO concentrations of either 2.5 or 0.5 mg/l. The lower DO concentration was set to limit nitrification during the study. At the higher DO concentration, the system was operated with nitrification occurring.

Ekama *et al.* (80) state that biological phosphorus removal will be adversely affected in biological combined nitrogen and phosphorus removal systems unless the DO concentration in the aerobic zone remains 1.5-3.0 mg/l. If the DO is too low, they claim that phosphorus removal may be reduced, nitrification will be limited, and a poor settling sludge may be developed. If too high, denitrification performance could be limited due to the increase in DO recycled to the first anoxic zone. A resultant higher nitrate nitrogen concentration could then affect the phosphorus release performance of the anaerobic zone.

3.3.5.9 Anaerobic Fermentator Zone Considerations

The anaerobic zone contact time for Modified Bardenpho and A/O systems has ranged from 0.9 hour for the Largo A/O facility to 2.0 hours for Modified Bardenpho facilities at the Payson and Kelowna plants. Early full-scale plant investigations at Palmetto, Florida, found that increasing the anaerobic detention time from 1.1 to 2.6 hours increased the percent total phosphorus removal from 59 to 71 percent (16). In a Bardenpho pilot-plant study, McLaren and Wood (81) found that the effluent soluble phosphorus concentration decreased from 3 to less than 1 mg/l as the anaerobic detention time was doubled from 2 to 4 hours. However, after establishing removal at the 4-hour detention time, an effluent soluble phosphorus concentration of less than 1 mg/l was maintained under variable anaerobic detention times. During investigations using an A/O pilot plant at the Saint Mars La Jaille, France wastewater facility, the anaerobic contact zone mixers were periodically turned off during the day and improved phosphorus removal was reported (75). With the mixers off, improved performance was attributed to a greater SDT. In these cases, it appears that the longer contact time results in the fermentation of particulates or materials that are more slowly converted to fatty acids. The necessity for and success of longer anaerobic contact times may vary depending on the strength and nature of the wastewater. Methods to increase the production of fatty acids to improve biological phosphorus removal performance will be discussed in Section 3.6.

Another important aspect of anaerobic contactor design and performance is to limit the amount of oxygen entering the zone. Any DO present will deplete readily-available substrate and thus reduce the amount of fatty acids that will be produced for biological phosphorus removal. The presence of excess DO was identified as causing poor performance for biological phosphorus removal at a number of full-scale South African facilities (57). This was also suspected, in combination with a weak wastewater, of causing poor phosphorus removal and filamentous sludge growth during a portion of the operating period of a U.S. Modified Bardenpho system (65). Possible sources of high DO input to the anaerobic zone have included high influent DO concentrations in the wastewater associated with infiltration, the use of Archimedes screw pumps for the return sludge or influent feed, cascading of wastewater through the influent channel flow measurement or grit removal systems, and vortices created by stirrers in the anaerobic basins. Such conditions should be avoided as much as possible in the design of biological phosphorus removal systems using the anaerobic fermentation step.

Anaerobic fermentation zones have been designed as single-stage, complete mixed basins or three to four

basins in series as in the A/O process designs. Experiments by Ekama *et al.* (30) support a multiple-stage design. They claimed improved phosphorus release and improved phosphorus removal for a multiple-stage system compared to a single-stage fermentation reactor operation. They explain this difference by a model that describes fermentation of readily-available substrate as a first-order reaction.

3.4 Equipment Requirements

Three major areas of equipment requirements for the Phostrip system are the stripper tank, the lime feed system, and the chemical precipitation tank. Piping and necessary pumping designs are also required to route a portion of the return sludge to the stripper, to provide elutriant to the stripper, to transport stripper underflow sludge to the aeration basin, to convey stripper overflow to the chemical treatment unit, and to feed lime to the chemical treatment step.

Stripper tanks are typically sludge thickening tanks with modifications for sludge inventory control and elutriation. The tank has a center well for sludge feeding, a scum baffle and overflow weir, a sludge rake mechanism, and sludge blanket level indicators. Underflow solids density probes have also been recommended by the Phostrip process supplier.

The lime feed system will normally include a lime storage tank, a slaking operation, and a lime slurry feed and control system. Mixing of the lime with the stripper supernatant may be accomplished by using static in-line mixers or a flash mix chamber with a mechanical stirrer. The flash mix chamber should have about a 1-minute detention time.

The chemical treatment units in the Phostrip process are typically solids contact units. In these units, relatively large, cone-shaped skirts form mixing zones in the center of the circular tanks to promote flocculation. With lime treatment, previously precipitated solids provide a seed for newly formed precipitates and floc growth. Heavier floc solids that fall out of the mixed zone settle out in the clarification area below and around the skirt. A raking mechanism is used to move settled solids to a center withdrawal point.

Equipment requirements for the mainstream biological phosphorus removal system are minimal and relatively simple. Mixers are needed to suspend the mixed liquor solids in the anaerobic and/or anoxic zones of the various designs. Such mixers have typically been designed with an energy input per unit volume of about 10 W/m³ (0.4 hp/1,000 ft³). Lower values than this may be desirable to minimize induced air entrainment by the mixers (57). Anti-vortex baffles are also used for this purpose. Internal recycle of sludge to anoxic or anaerobic zones is accomplished with low head, high capacity pumps.

3.5 Design Methodology

3.5.1 Phostrip Process

The major design considerations for the Phostrip process are the size of the stripper and solids contact tanks and the lime feed rate. The size of the solids contact tank will be a function of the stripper tank supernatant overflow rate. This will be determined by the return sludge feed rate to the stripper, the degree of solids thickening achieved, and the elutriation rate if the elutriant is composed of an outside flow instead of recycled stripper sludge. The lime feed rate will be affected by the stripper tank supernatant characteristics, which impact the ability to raise the pH for phosphorus precipitation, as well as the stripper tank supernatant overflow rate. Typical values for stripper and reactor-clarifier design were given in Table 3-2.

The stripper design procedure involves the following steps:

1. Determine or select the amount of return sludge that will pass through the stripper.
2. Select the stripper underflow sludge concentration.
3. Select the stripper SDT.
4. Based on the above, calculate the volume of sludge necessary in the stripper.
5. Using a solids flux analysis (82) or appropriate solids loadings, calculate the stripper area requirements.
6. Using information from steps 4 and 5, determine the sludge depth in the stripper.
7. Provide a selected supernatant water depth to obtain the total stripper sidewater depth. A supernatant water depth of 1.5 m (5 ft) has been recommended (77). The stripper depth may be increased to provide additional sludge inventory and operating flexibility.

The amount of the return sludge passing through the stripper is usually selected based on pilot-plant work or previous full-scale plant operating information. Another approach, presented by Peirano *et al.* (77), was developed from plant-scale Phostrip performance testing at Reno-Sparks. The phosphorus removal efficiency was correlated with three main operating parameters: the amount of return sludge passing through the stripper relative to the plant flow, the stripper SDT, and the stripper supernatant flow. The correlation developed can be expressed as follows:

$$1.85 - [\log (100 - E)]/2.11 = (SL \times D)^{1/2} (SU) \quad (3-1)$$

where,

- E = percent phosphorus removal
SL = return sludge passing through stripper tank,
100 lb dry solids/mil gal of system influent
flow
D = SDT, hr
SU = stripper supernatant flow as ratio of influent
flow

This relationship indicates that phosphorus removal is affected by the solids loading to the stripper and the stripper SDT. The following example illustrates the design procedure.

Wastewater and Plant Design Assumptions:

Influent flow = 10,000 m³/d (2.6 mgd)
Primary effluent TBOD = 120 mg/l
Primary effluent TP = 8 mg/l
Activated sludge recycle flow rate as percentage
of influent flow rate = 80 percent
Activated sludge recycle solids
concentration = 6,000 mg/l

Stripper Design Assumptions:

SDT = 10 hr
Underflow solids concentration = 9,000 mg/l
Return sludge flow rate to stripper
as percentage of influent flow rate = 25 percent

Design Steps:

1. Return sludge flow rate to stripper:

$$0.25 (10,000 \text{ m}^3/\text{d}) = 2,500 \text{ m}^3/\text{d}$$

Recycle sludge passed through stripper:

$$(0.25/0.8) \times 100 = 31 \text{ percent} \quad (\text{based on total return sludge flow rate})$$

2. Stripper underflow solids concentration = 9,000 mg/l

3. Stripper SDT = 10 hr

4. Stripper sludge volume produced/day (i.e., stripper underflow rate):

$$(2,500 \text{ m}^3/\text{d}) (6,000 \text{ mg/l} \div 9,000 \text{ mg/l}) = 1,667 \text{ m}^3/\text{d}$$

Net stripper tank sludge volume:

$$(1,667 \text{ m}^3/\text{d} \div 0.8) (10 \text{ hr}) (d/24 \text{ hr}) = 868 \text{ m}^3 (30,650 \text{ ft}^3)$$

This net estimate of stripper tank sludge volume required is based on the stripper underflow solids rate (or concentration) and an assumed density factor of 0.8 to account for the possibility of a lower thickened sludge concentration due to variations in the stripper operation.

5. Solids loading to stripper:

$$(6,000 \text{ mg/l}) (2,500 \text{ m}^3/\text{d}) (0.001 \text{ kg/mg}) = 15,000 \text{ kg/d} (33,070 \text{ lb/d})$$

Assume allowable solids flux rate for 9,000-mg/l underflow solids concentration = 50 kg/m²/d

$$\text{Stripper area req'd} = 15,000 \text{ kg/d} \div 50 \text{ kg/m}^2/\text{d} = 300 \text{ m}^2 (3,230 \text{ ft}^2)$$

$$\text{Overflow rate} = (2,500 \text{ m}^3/\text{d} \div 300 \text{ m}^2) (d/24 \text{ hr}) = 0.35 \text{ m/hr} (205 \text{ gpd/ft}^2)$$

6. Stripper sludge depth:

$$868 \text{ m}^3 \div 300 \text{ m}^2 = 2.9 \text{ m} (9.6 \text{ ft})$$

7. Minimum stripper depth:

$$1.5 \text{ m} + 2.9 \text{ m} = 4.4 \text{ m} (14 \text{ ft.})$$

[use total stripper depth of 5.5 m (18 ft) for added inventory flexibility]

8. Supernatant flow assuming primary effluent elutriation at 50 percent of stripper feed flow:

$$(2,500 \text{ m}^3/\text{d}) (0.50) + (2,500 \text{ m}^3/\text{d}) [1 - (6,000 \text{ mg/l} \div 9,000 \text{ mg/l})] = 2,083 \text{ m}^3/\text{d}$$

9. Solids contact unit for lime precipitation:

$$\begin{aligned} \text{Assume: Overflow rate} &= 49 \text{ m}^3/\text{m}^2/\text{d} \\ \text{Area} &= 2,083 \text{ m}^3/\text{d} \div 49 \text{ m}^3/\text{m}^2/\text{d} = 42.5 \text{ m}^2 \\ \text{Diameter} &= 7.4 \text{ m} (24 \text{ ft}) \end{aligned}$$

Lime feed at dose of 200 mg/l:

$$(2,083 \text{ m}^3/\text{d}) (200 \text{ mg/l}) (0.001) = 417 \text{ kg/d} (919 \text{ lb/d})$$

10. Check phosphorus removal; use phosphorus release in stripper of 0.01 g P/g VSS (see Table 3-2); assume 70 percent volatile solids:

$$(15,000 \text{ kg/d}) (0.70) (0.01 \text{ g/g}) = 105 \text{ kg/d} (236 \text{ lb/d})$$

Phosphorus removed by stripper supernatant treatment:

$$(105 \text{ kg/d} \times 2,083 \text{ m}^3/\text{d}) \div [2,500 \text{ m}^3/\text{d} + 2,500 \text{ m}^3/\text{d} (0.5)] = 58.3 \text{ kg/d} (126 \text{ lb/d})$$

11. Determine phosphorus content of waste sludge:

Total phosphorus in influent to activated sludge system:

$$(10,000 \text{ m}^3/\text{d})(8 \text{ mg/l}) (0.001) = 80 \text{ kg/d} (176 \text{ lb/d})$$

Assume TP in effluent = 0.5 mg/l

Phosphorus in activated sludge waste solids:

$$(80 - 58.3) \text{ kg/d} - 0.5 \text{ mg/l} (10,000 \text{ m}^3/\text{d}) (0.001) = 16.7 \text{ kg/d} (37 \text{ lb/d})$$

Assume net sludge yield for biological system following primary treatment = 0.55 g TSS/g TBOD removed

$$\text{TBOD removed} = 120 \text{ mg/l} - 10 \text{ mg/l} = 110 \text{ mg/l}$$

Net sludge produced

$$= (110 \text{ mg/l}) (0.55 \text{ g/g}) (10,000 \text{ m}^3/\text{d}) (0.001) = 605 \text{ kg/d} (1,334 \text{ lb/d})$$

P in waste sludge

$$= (16.7 \text{ kg/d} \div 605 \text{ kg/d}) \times 100 = 2.8 \text{ percent}$$

This is a relatively low sludge phosphorus content compared to mainstream biological phosphorus removal systems.

3.5.2 Mainstream Biological Phosphorus Removal Processes

A variety of process configurations have been presented for mainstream biological phosphorus removal. While the aerobic zones may be designed for different treatment objectives or the internal recycle and nitrate reduction schemes may be different, there are common design considerations that apply to all of these systems. These considerations include the design of the anaerobic zone, the need for sufficient time and DO in the aerobic zone, the denitrification reactor design when needed, and sludge handling. Another major consideration is the effluent phosphorus level that can be achieved and whether chemical addition and/or effluent filtration is necessary to meet required treatment levels. In many cases, pilot-plant or bench-scale studies should be recommended to determine the final design since treatment performance is very sensitive to individual wastewater characteristics.

The anaerobic zone contact time is presently based on pilot-plant studies or previous experience and has ranged from 0.9 to 2.0 hours. Staging of the

anaerobic zone should theoretically decrease the required detention time for the fermentation of soluble organics. This has to be balanced against the higher cost for an increased number of mixers and the use of more divider walls. A DO concentration of greater than 2.0 mg/l is commonly recommended in the aerobic zone. The size of the aerobic zone should be kept as small as is consistent with the overall treatment objective of maximizing phosphorus removal. Sufficient aerobic time is needed for phosphorus uptake.

Waste sludge from biological phosphorus removal systems is handled in a manner to minimize any recycle of released phosphorus back to the activated sludge system. As gravity thickening normally results in substantial release of phosphorus from the sludge, dissolved air flotation thickening has been used where sludge thickening is needed. Provisions may be required to chemically treat any recycle streams from digestion of the phosphorus laden sludge. As solids are lysed and destroyed in aerobic digestion, a proportional amount of phosphorus is released to the liquor. The same phenomenon is expected with anaerobic digestion, but the Pontiac, Michigan study (67) did not note significant levels of phosphorus in the digester supernatant. An ammonium-magnesium-phosphate precipitate may have formed in the digester, and further studies are needed in this area.

3.5.2.1 Sludge Production

No significant differences in sludge production for biological phosphorus removal systems compared to typical sludge yield values are reported in the literature for the operating conditions employed. However, if mixed liquor solids are capable of storing phosphorus, some increase in net sludge yield should be expected. To calculate the increase, an estimate of the mass of the associated chemical constituents is needed. This may be approximated from the constituents reported in solution during phosphorus release (Section 3.1.1). Table 3-14 summarizes the total expected stored mass/unit of phosphorus stored.

The following example illustrates the increase in sludge mass associated with biological phosphorus removal:

Assume:

Net solids yield = 0.70 g TSS/g TBOD removed
Normal phosphorus content = 2 percent
Biological phosphorus removal raises the phosphorus content in the dry solids to 4 percent

Calculate:

$$\begin{aligned} \text{Solids yield/100 mg TBOD removed} &= 70 \text{ mg} \\ \text{Normal P removed/100 mg TBOD removed} &= 0.02 (70) = 1.4 \text{ mg} \end{aligned}$$

Table 3-14. Approximated Mass of Phosphorus Storage Compounds.

Constituent	Molecular Weight	Mole per Mole of P	g/g P
Mg	24.3	0.28	0.22
K	39.1	0.20	0.25
Ca	40	0.09	0.12
O	16	4	2.06
P	31	1	1.0
Total			3.65

P_B = additional P removed by biological phosphorus removal

$$(1.4 + P_B) \div (70 + 3.65 P_B) = 0.04$$

$P_B = 1.64 \text{ mg}/100 \text{ mg TBOD removed}$

Amount of sludge produced with biological phosphorus removal/100 mg TBOD removed:

$$70 + 3.65 (1.64) = 76 \text{ mg}$$

Ratio of biological phosphorus process sludge production to conventional process sludge production:

$$76/70 = 1.085$$

The sludge yield increased by 8.5 percent for this example. If the phosphorus content of the waste sludge increased to 5 percent, the net sludge yield increase is estimated to be 13 percent. Thus, an increase in the mass of waste solids is expected for biological phosphorus removal systems. However, the overall impact on sludge handling may not be negative due to the excellent sludge thickening and dewatering characteristics reported for these systems. Mixed liquor SVI values generally less than 80 ml/g have been reported for Modified Bardenpho and A/O systems (12,16,29).

3.5.2.2 Phosphorus Removal Efficiency

In the absence of bench-scale or pilot-plant studies, estimates of biological phosphorus removal efficiency can be made to determine if chemical addition and/or filtration may be necessary to meet the effluent requirement. The selection of filtration during the design phase may also be affected by the assumed secondary clarification efficiency. If an effluent TSS concentration of less than 10-12 mg/l must be attained to achieve an effluent total phosphorus concentration of less than 1 mg/l, filtration would generally be required.

The amount of phosphorus that may be removed in a biological phosphorus removal system will be a

function of the net sludge production, the phosphorus content of the sludge, and the amount of BOD removed. This is shown as follows:

$$(Y_n) (F_p) = DP/DBOD \quad (3-2)$$

where,

Y_n = net solids yield, g TSS/g TBOD removed

F_p = fraction of P in dry solids, g P/g TSS

$DP/DBOD$ = total phosphorus removed/unit of TBOD removed, g TP/g TBOD

The net solids yield is a function of the system operating SRT value and influent wastewater characteristics. The use of primary treatment will also lower the net solids yield since much of the influent inert solids will be removed during primary settling. The fraction of phosphorus in the solids has been shown to be quite variable depending on influent wastewater characteristics and operating conditions. A value for F_p can be selected based on results reported for other facilities. Figure 3-12 shows the TBOD:TP removal ratio as a function of SRT and an assumed F_p value of 0.05 for a system with no primary treatment (83). The net solids yield as a function of SRT was taken from a curve given for municipal wastewaters in WPCF MOP 8 (84). A system with primary treatment would have a lower Y_n but a higher F_p value since less inert material would be contained in the mixed liquor solids.

Example:

Assume:

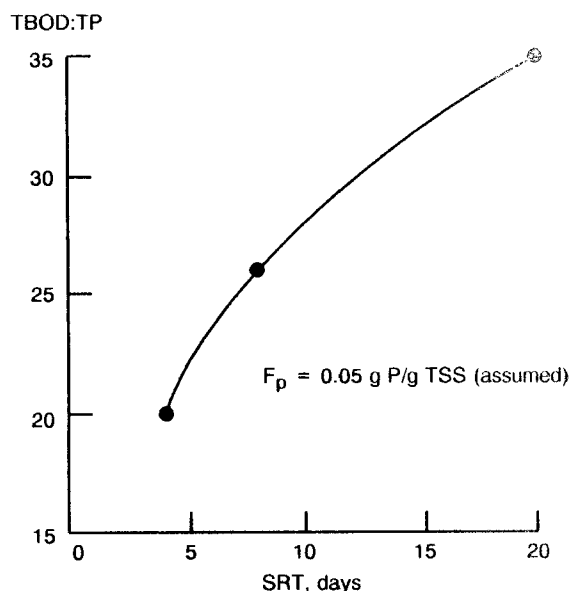
$F_p = 0.05 \text{ g P/g TSS}$
 Influent TBOD = 160 mg/l
 Influent TP = 7.5 mg/l
 Influent TBOD:TP = 21.3
 Effluent TSS = 12 mg/l
 Effluent TBOD = 5 mg/l
 No primary treatment
 $T = 20^\circ\text{C}$

Calculate:

Design SRT, days	5	10	20
Y_n (MOP 8), g/g	0.92	0.81	0.70
$DP/DBOD$ (Eq. 3-2)	0.046	0.041	0.035
DBOD, mg/l	155	155	155
DP, mg/l	7.1	6.4	5.4
Eff. SP			
(7.5 - DP), mg/l	0.4	1.1	2.1
Eff. part. P, mg/l	0.6	0.6	0.6
Eff. TP, mg/l	1.0	1.7	2.7
TP removal, percent	87	77	64

The above example shows the effect of SRT on estimated total phosphorus removal efficiency. If an

Figure 3-12. TBOD:TP removal vs. solids retention time (SRT).



effluent total phosphorus concentration of 1.0 mg/l is required, the 5-day SRT operation may meet performance standards without chemical addition or filtration. The longer SRT operating systems would require chemical addition to further reduce effluent soluble phosphorus concentrations. The example is based on a relatively low influent TBOD:TP ratio compared to some of the ratios observed in the plant performance section. However, the ratio used is within the range of values that can be expected for some domestic wastewaters (82).

The effect of nitrate nitrogen can be estimated by calculating the reduction in available BOD due to denitrification.

Assume:

Effective nitrate nitrogen concentration to anaerobic zone = 5 mg/l after mixing of recycle sludge and influent.

TBOD consumed for denitrification:

$$(4 \text{ mg TBOD/mg NO}_3\text{-N}) (5 \text{ mg/l}) = 20 \text{ mg/l}$$

For SRT = 5 days, remaining influent TBOD (i.e., TBOD available for biological phosphorus removal)
 $= 160 \text{ mg/l} - 20 \text{ mg/l} = 140 \text{ mg/l}$

$$\text{DBOD} = 140 \text{ mg/l} - 5 \text{ mg/l} = 135 \text{ mg/l}$$

$$\text{DP} = 0.046 (135 \text{ mg/l}) = 6.2 \text{ mg/l}$$

$$\begin{aligned} \text{Effluent SP} &= 7.5 \text{ mg/l} - 6.2 \text{ mg/l} \\ &= 1.3 \text{ mg/l vs. } 0.4 \text{ mg/l previously} \end{aligned}$$

Thus, for the low influent TBOD:TP ratio used for this example, nitrification and the presence of nitrates in the return sludge could significantly affect the effluent soluble and total phosphorus concentrations.

3.5.2.3 Ultrate Nitrogen Removal Design

Figure 3-13 illustrates the two modes of denitrification operation used in biological phosphorus removal systems. Nitrified mixed liquor is recycled to a pre-denitrification zone in the Modified Bardenpho process and also in the A/O process when nitrification occurs. The recycle ratio is generally in the range of 4:1 based on the influent flow. In this zone, the incoming substrate drives the denitrification reaction as the facultative organisms use nitrate-released oxygen as the electron acceptor in lieu of DO. The oxygen equivalent of the nitrate radical is 2.86 g O₂/g NO₃-N. The Modified Bardenpho process has a second anoxic tank, or post-denitrification zone, in addition to the pre-denitrification zone. In the second anoxic zone, the denitrification rate is driven by the endogenous respiration oxygen demand of the mixed liquor since the influent substrate is depleted after the nitrification step.

The design objectives for biological phosphorus removal systems incorporating denitrification are to first determine the amount of nitrate nitrogen entering the pre-denitrification and post-denitrification zones and then to determine the volume of the anoxic zones. A critical design aspect is the mixed liquor denitrification rate occurring in each type of anoxic zone. A design approach for denitrification will be briefly presented here, since nitrate reduction can be an important consideration in biological phosphorus removal systems.

The first step in the design is the preparation of a mass balance to determine the amount of influent nitrogen that will be oxidized to nitrate nitrogen. It is generally assumed that the distribution of influent nitrogen is to nitrate nitrogen, effluent ammonium nitrogen, and solids synthesis:

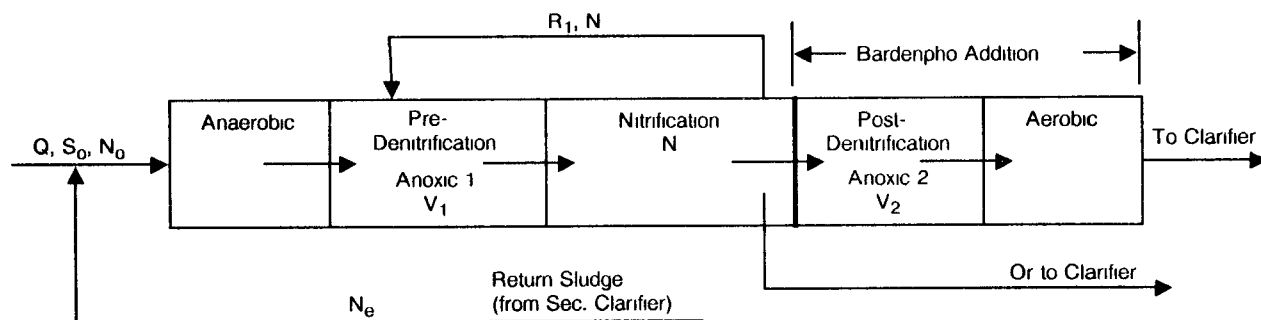
$$\text{NO} = \text{N}_0 - \text{NH}_e - \text{N}_{\text{syn}} \quad (3-3)$$

where,

NO = amount of influent nitrogen converted to oxidized nitrogen, mg/l
 N_0 = influent total nitrogen, mg/l
 NH_e = effluent ammonium nitrogen, mg/l
 N_{syn} = amount of influent nitrogen used in solids synthesis, mg/l

The amount of nitrogen used in synthesis can be estimated from the amount of BOD removed, the net solids yield as a function of SRT, and the nitrogen content of the mixed liquor. The nitrogen content of bacteria is 10-12 percent, but a lower value will typically be measured for mixed liquor solids because

Figure 3-13. Pre-denitrification and post-denitrification schemes in biological phosphorus removal systems.



of the presence of inerts and non-biological solids. Values of 5-8 percent may be more appropriate:

$$N_{syn} = Y_n (DBOD) F_n \quad (3-4)$$

where,

F_n = fraction of nitrogen in mixed liquor solids, g/g

Once NO is determined, the next step is to perform a mass balance describing the distribution of the nitrate produced in the nitrification zone, which results in the following:

$$N = NO / (R + r + 1) \quad (3-5)$$

where,

N = nitrate nitrogen concentration in the nitrification zone, mg/l
 R = ratio of internal recycle flow (to the pre-denitrification zone) to influent flow
 r = ratio of return sludge flow to influent flow

Equation 3-5 is applicable to both A/O and Modified Bardenpho system designs. The rate of nitrate nitrogen addition to either denitrification zone can be calculated once the value of N is determined. The design approach assumes that all of the nitrate nitrogen entering the anoxic zones is completely reduced even though a residual nitrate nitrogen concentration of 0.3-0.5 mg/l may exist. The volume of the denitrification zones is then determined based on the amount of nitrate nitrogen entering the zone and the specific denitrification rate:

Anoxic 1 Volume (applies to both A/O and Modified Bardenpho):

$$V_1 = RQN / [(X) (SDNR_1)] \quad (3-6)$$

Anoxic 2 Volume (Modified Bardenpho)

$$V_2 = [(1 + r) NQ] / [(X) (SDNR_2)] \quad (3-7)$$

where,

V_1 = volume of pre-denitrification zone, m^3
 V_2 = volume of post-denitrification zone, m^3
 Q = influent flow, m^3/d
 X = MLSS concentration, mg/l
 $SDNR_1$ = specific denitrification rate in pre-denitrification zone, g NO_3-N/g X/d .
 $SDNR_2$ = specific denitrification rate in post-denitrification zone, g NO_3-N/g X/d .

The $SDNR$ has been predicted from the specific oxygen uptake rate as follows:

$$SDNR = F_d \text{ SOUR} / 2.86 \quad (3-8)$$

where,

F_d = fraction of substrate reaction rate when nitrogen-released oxygen is the electron acceptor vs. when DO is the electron acceptor, g/g.
 $SOUR$ = specific oxygen uptake rate, g O_2/g TSS/d

Previous investigators have found values for F_d of 0.41-0.55 for systems in which the mixed liquor was subject to both aerobic and anoxic conditions (85-87). The reduced reaction rate during nitrate reduction is attributed to the possibility that the entire biological population cannot use nitrate as an electron acceptor in the absence of DO , and that the biological reaction rate may be slower when nitrate is the electron acceptor.

With this approach, good agreement has been observed for pre-denitrification $SDNR$ reaction rates predicted from treatment of a tannery wastewater in the presence of plentiful substrate (87) and post-denitrification $SDNR$ reaction rates predicted from endogenous respiration for domestic wastewater with limited available substrate (85). The $SDNR$ prediction for pre-denitrification with domestic wastewater has not been demonstrated and is limited by unknown reaction rates for particulate and soluble BOD. An

SDNR relationship based on the F/M loading to the pre-denitrification zone has been demonstrated for pilot- and full-scale plant data (64):

$$\text{SDNR}_1 = 0.03 (F/M)_1 + 0.029 \quad (3-9)$$

$$(F/M)_1 = Q S_0 / XV_1 \quad (3-10)$$

where,

$(F/M)_1$ = food-to-mass loading in pre-denitrification zone, g TBOD/g MLSS/d
 S_0 = influent TBOD, mg/l

The $(F/M)_1$ value can be calculated for staged or completely mixed pre-denitrification zone as a function of the volume (V_1) selected for the zone.

The SDNR for the post-denitrification zone can be calculated as follows using an F_d factor equal to 0.5:

$$\text{SDNR}_2 = F_d A_n \div [(2.86 Y_n)(\text{SRT})] \quad (3-11)$$

$$\text{SDNR}_2 = 0.175 (A_n/Y_n)(1/\text{SRT}) \quad (3-12)$$

where,

A_n = net amount of oxygen required per unit of TBOD removed, g O_2 /g DBOD

Using values for A_n and Y_n as a function of SRT from the WPCF MOP 8 publication (84), the SDNR_2 relationship shown in Figure 3-14 was developed. Bench- and full-scale results treating domestic wastewater compared well with the curve. Different curves would be developed based on changes in A_n and Y_n due to the system SRT, wastewater temperature, wastewater characteristics, and operation with or without primary treatment.

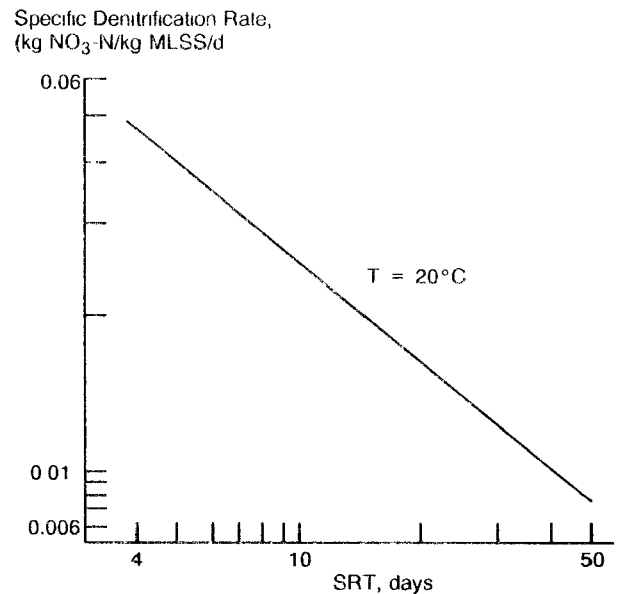
Design Example:

Calculate the anoxic zone detention times for the Modified Bardenpho process.

Assume:

No primary treatment
 $Q = 10,000 \text{ m}^3/\text{d}$
 $N_0 = 30 \text{ mg/l}$
 $\text{TBOD} = 200 \text{ mg/l}$
 $\text{DBOD} = 190 \text{ mg/l}$
 $T = 20^\circ\text{C}$
 $\text{Design SRT} = 20 \text{ days}$
 $Y_n = 0.70 \text{ g TSS/g DBOD}$
 $A_n = 1.25 \text{ g } O_2/\text{g DBOD}$
 $F_n = 0.06 \text{ g N/g MLSS}$
 $\text{NH}_e = 0.9 \text{ mg/l}$
 $X = 3,500 \text{ mg/l}$
 $R = 4.0$
 $r = 1.0$
 Completely mixed denitrification zones

Figure 3-14. Specific denitrification rate.



The actual design SRT and SDNR values may be determined using an iterative procedure.

Based on the nitrification and denitrification rates, a total system volume and SRT is calculated. This may then change SDNR_2 , which is then used to calculate a new system volume and SRT.

Step 1: Determine nitrate nitrogen concentration in the nitrification zone.

$$N_{\text{syn}} = 0.70 \text{ g/g } (190 \text{ mg/l})(0.06 \text{ g/g}) = 8.0 \text{ mg/l}$$

$$\text{NO} = 30 - 0.9 - 8.0 = 21.1 \text{ mg/l}$$

$$N = 21.1 \div (4 + 1 + 1) = 3.5 \text{ mg/l NO}_3\text{-N}$$

Step 2: Determine V_1 using Equations 3-6, 3-9, and 3-10.

$$V_1 = [(4)(10,000 \text{ m}^3/\text{d})(3.5 \text{ mg/l})] \div [(3,500 \text{ mg/l})(\text{SDNR}_1)] = 40/\text{SDNR}_1$$

$$(F/M)_1 = [(10,000 \text{ m}^3/\text{d})(200 \text{ mg/l})] \div [(3,500 \text{ mg/l})(V_1)] = 571.4/V_1$$

$$\text{Assume: } V_1 = 1,000 \text{ m}^3$$

$$\text{Assumed } \text{SDNR}_1 = 40 \div 1,000 = 0.04 \text{ g NO}_3\text{-N/g TSS/d}$$

$$(F/M)_1 = 571.4 \div 1,000 = 0.57 \text{ g TBOD/g TSS/d}$$

$$\begin{aligned}\text{Predicted SDNR}_1 &= 0.03 (0.57) + 0.029 \\ &= 0.046 \text{ g NO}_3\text{-N/g TSS/d}\end{aligned}$$

OK - Assumed SDNR_1 lower than predicted

$$\text{Try: } V_1 = 800 \text{ m}^3$$

$$\begin{aligned}\text{Assumed SDNR}_1 &= 40 \div 800 \\ &= 0.05 \text{ g NO}_3\text{-N/g TSS/d}\end{aligned}$$

$$\begin{aligned}(F/M)_1 &= 571.4 \div 800 \\ &= 0.71 \text{ g TBOD/g MLSS/d}\end{aligned}$$

$$\begin{aligned}\text{Predicted SDNR}_1 &= 0.03 (0.71) + 0.029 \\ &= 0.05 \text{ g NO}_3\text{-N/g TSS/d} \\ &\quad - \text{ OK}\end{aligned}$$

$$\begin{aligned}V_1/Q &= (800 \text{ m}^3 \div 10,000 \text{ m}^3/\text{d})(24 \text{ hr/d}) \\ &= 1.92 \text{ hr}\end{aligned}$$

Step 3: Determine V_2 using Equations 3-7 and 3-12.

$$\begin{aligned}V_2 &= [(1 + 1)(3.5 \text{ mg/l})(10,000 \text{ m}^3/\text{d})] \\ &\quad \div [(3,500 \text{ mg/l})(\text{SDNR}_2)] \\ &= 20/\text{SDNR}_2\end{aligned}$$

$$\begin{aligned}\text{SDNR}_2 &= 0.175 (1.25 \text{ g/g} \div 0.70 \text{ g/g})(1/20 \text{ d}) \\ &= 0.0156 \text{ g NO}_3\text{-N/g TSS/d}\end{aligned}$$

$$V_2 = 20 \div 0.0156 = 1,280 \text{ m}^3$$

$$\begin{aligned}V_2/Q &= (1,280 \text{ m}^3 \div 10,000 \text{ m}^3/\text{d})(24 \text{ hr/d}) \\ &= 3.1 \text{ hr}\end{aligned}$$

The design procedure should also check the TBOD:NO ratio to determine that there is sufficient TBOD available for the amount of nitrate nitrogen to be reduced. A ratio of at least 4:1 is recommended. In this case, there is sufficient TBOD available for denitrification.

3.6 Process Modifications to Improve Performance

The major performance limitation of biological phosphorus removal systems is the amount of volatile fatty acids available relative to the amount of phosphorus that must be removed by the phosphorus-storing microorganisms. As described in Section 3.3.5.3, wastewaters with lower influent BOD-to-phosphorus ratios may not produce a sufficient level of volatile fatty acids (VFAs) in the fermentation zone to fully trigger the biological phosphorus removal mechanisms. In some installations, it has been necessary to add chemicals to reduce the effluent total phosphorus concentration to discharge requirement levels. Another means of improving performance is to increase the amount of VFAs available to the microorganisms.

It is generally accepted that the major contribution of VFAs produced in mainstream biological phosphorus removal systems is from readily-degradable soluble BOD entering the fermentation zone. For many wastewaters this may only represent 30-60 percent of the influent TBOD. Osborn and Nicholls (15), suspecting that VFAs were important substrates for biological phosphorus removal, operated a primary sludge treatment digester at a high loading to encourage only acid fermentation. The fermented sludge was then fed to the anaerobic zone of a Modified Bardenpho system. The phosphorus removal was excellent, but the sludge addition increased the aeration energy requirements of the plant. Eventually methane fermentation developed in the digester and phosphorus removal efficiency declined.

Oldham and Stevens (62) presented data showing the benefits of using primary sludge fermentation products at the Kelowna Modified Bardenpho facility. The primary sludge at this facility is directed to a gravity thickener where it is held long enough for acid fermentation to develop. The thickened sludge was passed through a 2.5-mm screen, and the screened liquid containing fine solids was directed to the plant fermentation zones. The fermentation stream was alternately directed to both and one of the two modules to compare performance with and without the fermentation product addition. When added to both modules, the effluent soluble phosphorus concentration was generally below 0.5 mg/l. During the alternating operation, the module receiving the fermenter liquor immediately exhibited a higher degree of phosphorus release in the fermentation zone and effluent soluble phosphorus concentrations well below 0.5 mg/l within about 3 days. The module not receiving the fermenter liquor produced effluent soluble phosphorus concentrations generally between 2 and 3 mg/l. During these tests, the VFA concentration of the thickener liquor was 110-140 mg/l. Since the fermenter liquor flow rate was only 8-10 percent of the influent flow rate, the increased VFA concentration in the influent was 9-10 mg/l. The sludge depth of the thickener was later increased to promote additional solids detention time and VFA production. The VFA concentration of the fermenter liquor increased to 200-300 mg/l, but this was followed by a decline in phosphorus removal efficiency. The pH of the thickener liquor also decreased. Barnard (88) postulated that the lower pH resulted in fermentation products that were not readily available to the phosphorus-storing microorganisms. Rabinowitz and Oldham (89) carried out UCT pilot-plant studies that were also fed settled supernatant from primary sludge fermentation. The sludge was fermented in a two-stage, completely mixed reactor followed by a clarifier for solids separation and sludge return to the first-stage sludge fermenter. The settled liquid contained VFAs in the range of 150-185 mg/l. The phosphorus removal in two sets of experiments increased by 100 percent and 47 percent

after the fermented liquid addition. The VFA production averaged 0.09 mg/mg COD applied to the fermenter.

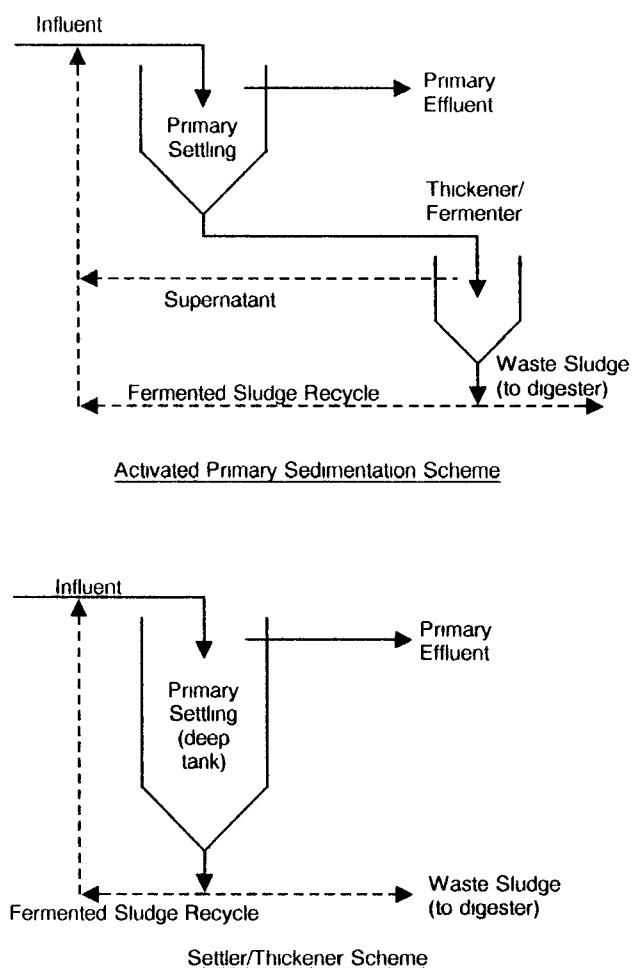
Figure 3-15 shows possible design schemes for primary sludge fermentation. The first is termed the activated primary sedimentation tank by Barnard (90). The recycling of thickened fermented solids serves a number of purposes. First, it provides mixing of the newly settled solids with the fermentation organisms. The acids produced in the thickener are also elutriated in the primary tank and then directed to the activated sludge process. In this way, primary treatment is also used to decrease the loading to the secondary treatment step and the size of the activated sludge system. Another advantage claimed by Barnard is that the pH of the solids in the fermenter is better buffered, resulting in the production of the preferred VFAs. The settler/thickener (deep tank) design can accomplish the same objectives as the activated primary sedimentation tank. Deep tank designs have been used previously to provide both settling and thickening. Rabinowitz (34) proposed a design similar to the activated primary sedimentation tank, but the pumping rate out of the thickener is controlled to maintain the desired sludge detention time for fermentation.

3.7 Retrofit Considerations

The sidestream treatment feature of the Phostrip process makes it readily adaptable for retrofitting existing facilities. Separate tankage and piping are added to strip phosphorus from a portion of the return sludge, return the stripped sludge to the activated sludge system, and lime treat the supernatant from the stripper tank. The design features for retrofitting are thus similar to the design aspects discussed for a new facility. The source of the elutriant flow and the organic loading to the activated sludge system are important design considerations. An elutriant source that is high in nitrates will require a longer stripper solids detention time and could negatively impact stripper performance. Activated sludge systems operated at longer SRTs will have a less active sludge, which can impact the stripper detention time and/or performance. For example, the Phostrip process has not been applied to systems operating with an extended aeration sludge age. As discussed in Chapter 5, additional sludge resulting from lime treatment of the stripper supernatant must be considered.

For mainstream retrofit alternatives, the design choice could be an operationally modified activated sludge system, an A/O system, or a Modified Bardenpho system. In all three cases, an anaerobic fermentation zone must be provided at the head end of the activated sludge facility. The retrofit design involves determining the volume requirements of the anaerobic

Figure 3-15. Primary sludge fermentation design schemes.



zones and anoxic zones where used. Some or all of the additional volume requirements may be available in the existing plant tankage or they may have to be added. In the latter case, the hydraulic and physical arrangements of the specific plant will be a factor in selecting the most economical modification.

Excess tank volume may be available for retrofitting for the following reasons:

1. The plant is underloaded, and anticipated future loading increases are less than originally expected.
2. The better SVI associated with biological phosphorus removal will allow operation with a much higher MLSS concentration than the present operation.
3. The system operating SRT can be reduced from the original design without a loss in effluent quality. A lower SRT value has been shown to improve the performance of biological phosphorus removal processes.

In the case of the A/O process without nitrification, the retrofit requirement may require additional tank volume of only 45-minutes detention time. This small volume may frequently be available in the existing system, especially when the improved sludge thickening characteristics are considered.

For all of the mainstream processes, the retrofit design must consider the processing of the waste activated sludge and the potential release and recycle of phosphorus to the activated sludge system. Further work is needed to investigate the fate of released phosphorus in anaerobic digesters. Aerobic digestion will result in a phosphorus release that is proportional to the sludge mass reduction. Removal of the sludge for land application is a favored alternative that also takes advantage of the higher nutrient content of the waste sludge.

The choice of the retrofit system will depend on treatment objectives, wastewater characteristics, and economics. In all cases, the biological phosphorus removal retrofit design should be compared to the chemical treatment alternative for phosphorus removal. Compared to the biological methods, chemical alternatives will have a higher operating cost due to chemical addition and increased sludge handling. As discussed in Section 3.5.2.1, sludge production from a biological phosphorus removal system is expected to be only slightly greater than that from a conventional activated sludge system. The biological phosphorus removal alternatives may require a higher initial capital investment for facility modifications but, in the long term, a net savings may result from the savings in operating costs.

Retrofit economic comparisons are site specific. Numerous factors are involved, and it is difficult to make general statements concerning these comparisons. As shown at Pontiac, Michigan (67), it was extremely simple to modify the existing system to operate in the A/O process mode and the capital cost was minimal. The high influent BOD:P ratio also favored the selection of a mainstream biological phosphorus removal process. On the other hand, weak wastewaters with a low influent BOD:P ratio generally favor the selection of a chemical treatment alternative, the Phostrip process, or perhaps a mainstream biological phosphorus removal process coupled with primary sludge fermentation.

Treatment needs will also affect the retrofit process selection and design. If a high level of nitrogen removal in addition to BOD and phosphorus removal are required, the Modified Bardenpho process is a prime candidate. If a lesser degree of nitrogen removal is desired along with BOD and phosphorus removal, the A²/O and UCT processes should also be considered. If only nitrification and BOD removal are required, the A/O process with or without an anoxic zone, the UCT process, an operationally modified

activated sludge system, and the Phostrip process are all candidates. If BOD removal only is the objective, all of the above processes with the exception of the Modified Bardenpho, UCT, and A²/O processes should be considered.

An operationally modified activated sludge system may be considered a higher risk alternative for retrofitting since it generally lacks the well defined anaerobic-aerobic zones of the UCT, A/O, and Modified Bardenpho processes. However, with favorable wastewater characteristics and a relatively large anaerobic zone, such systems have been able to achieve effluent phosphorus concentrations equivalent to those of the staged systems. The plug flow systems reporting "luxury uptake" in the early literature also showed that operationally modified systems could achieve good phosphorus removal even though coarse bubble aeration was applied to the anaerobic fermentation zone.

Many plants can be easily modified to create fermentation zones. This can be done by turning off selected aerators, decreasing the air supply to sparged air headers at the head of the aeration tank, turning off aerators and adding mixers at the head of the aeration tank, or by recycling the return activated sludge through existing primary clarifiers for anaerobic contacting with raw wastewater. In the latter case, the entire primary solids inventory would then be directed to the activated sludge aeration tank. For a modified operation of this type, existing oxygen transfer and organic treatment capacities would have to be carefully evaluated.

Operationally modified activated sludge systems may also be used with chemical treatment phosphorus removal systems, where appropriate, to decrease chemical treatment costs. An advantage of the operationally modified activated sludge alternative is that it can usually be easily tested in the existing plant before final process design decisions are developed.

In summary, biological phosphorus removal process options generally are easily adaptable for plant retrofitting. The process choice, design, and economics, however, will be extremely site specific. All the design considerations described for new facilities apply also to retrofit designs.

3.8 Case Histories

3.8.1 Phostrip Process - Little Patuxent, Maryland

The Little Patuxent (Savage, Maryland) wastewater treatment plant has a design treatment capacity of 56,800 m³/d (15 mgd) and was started up in 1982. The plant treatment scheme includes primary treatment; a first-stage, high-rate activated sludge process; separate-stage nitrification; and chemical coagulation/flocculation with filtration for residual

phosphorus and suspended solids removal. Primary and waste biological sludges are aerobically digested before being gravity thickened with the waste chemical sludge. The thickened sludge is dewatered with a belt filter press before final disposal.

The Phostrip process is incorporated within the first-stage activated sludge system. Partial nitrification also occurs in the first-stage system. System operation and performance were studied in 1984 and 1985 as part of a U.S. EPA-sponsored evaluation of full-scale biological phosphorus removal installations. The information given here is from that study (48).

Operating changes were made in the first-stage activated sludge system in April 1985 to improve treatment performance. The first-stage activated sludge operation was changed from step feed to plug flow to improve phosphorus uptake. During the step-feed operating mode, the mixed liquor DO concentration was normally less than 2 mg/l and phosphorus release occurred in the secondary clarifiers. During these operating conditions, the effluent total phosphorus concentration averaged 2 mg/l.

Other changes made concurrently with the switch to a plug flow operating mode were to increase the mixed liquor DO concentration to 4 mg/l at the effluent end of the aeration basin to prevent phosphorus release in the secondary clarifiers and to increase the return sludge ratio to decrease the secondary clarifier blanket level. Other changes were also simultaneously made in the stripper operation. During April 1985, the first-stage activated sludge operating parameters were reported as follows:

F/M loading = 0.5 kg TBOD/kg MLVSS/d
MLSS = 2,000 mg/l
SRT = 4.6 days
HRT = 3.0 hr
Return sludge ratio = 0.57

The elutriation rate to the stripper was decreased in April 1985 from 121 to 50 percent of the stripper feed flow rate. Recycled stripper underflow was also used as an elutriant source in addition to the previously-used reactor-clarifier overflow. The SDT in the stripper was maintained at 7 hours. The return sludge flow to the stripper was increased from 22 to 34 percent of the system influent flow rate.

The orthophosphorous concentration of the stripper liquor increased from 7.2 to 17.6 mg/l after the above changes. As shown previously in Table 3-5, the average monthly effluent phosphorus concentration from the first-stage activated sludge system decreased from 1.7 to 0.5 mg/l.

The overflow rate of the reactor-clarifier used for chemical treatment of the stripper overflow was 34

m³/d/m² (840 gpd/ft²). The lime dosage was about 100 mg/l to maintain a pH of 9.5 for phosphorus precipitation. About 50-60 percent of the total amount of phosphorus removed in the system was removed in the stripper operation. The remainder was removed via the waste activated sludge. The phosphorus content of the waste solids on a dry weight basis was about 3.9 percent.

April 1985 treatment performance for the first-stage activated sludge system is summarized in Table 3-15. Total phosphorus removal efficiency averaged 94 percent in spite of a relatively weak influent TBOD concentration that resulted in an influent TBOD:TP ratio of 13.1. Effluent TBOD and TSS concentrations were also very good.

Table 3-15. Summary of Phostrip Process Treatment Performance for Little Patuxent Plant - April 1985.

Flow, m ³ /d	40,900
<u>Influent</u>	
Total P, mg/l	8.8
<u>1st-Stage AS Influent</u>	
TBOD, mg/l	92
Total P, mg/l	7.0
Ortho P, mg/l	4.8
TBOD:TP	13.1
<u>1st-Stage AS Effluent</u>	
TBOD, mg/l	6
TSS, mg/l	11
Total P, mg/l	0.5
Ortho P, mg/l	0.1
NH ₄ -N, mg/l	9.5
Total N	21

3.8.2 Modified Bardenpho Process - Kelowna, Canada

The city of Kelowna, Canada, is located in the Okanagan Valley in the Province of British Columbia. Okanagan Lake is generally considered to be in an oligotrophic state, and, in the late 1960s, algal blooms on several areas of the lake caused concern. A water basin study recommended 80 percent total phosphorus reduction from sewerage system discharge points. In addition, nitrogen was also considered limiting in certain areas of the lake. Several treatment schemes were considered including land application and advanced physical-chemical treatment before selecting the Modified Bardenpho process on the basis of lowest cost.

The existing facility had a design capacity of 11,400 m³/d (3.0 mgd) and consisted of primary treatment,

conventional activated sludge, and trickling filters for a portion of the flow. The existing inlet works, primary clarifiers, and secondary clarifiers were incorporated in the new plant. The upgraded plant design flow is 22,700 m³/d (6.0 mgd), with a maximum hourly flow of twice the average flow. Raw wastewater flows through conventional headworks consisting of a barminnutor, grit chambers, and Parshall flumes for flow measurement. The three existing primary clarifiers are now operated in a high-rate mode prior to the Modified Bardenpho system reactor.

During the design phase, the Modified Bardenpho process was in its early period of application and an effort was made to provide maximum operational flexibility. The process was divided into two modules with 21 square cells in each module. The design loadings have been given in Table 3-6 in Section 3.3.2. Following secondary clarification, polishing filters and chlorination are employed before effluent discharge. The primary sludge is directed to a gravity thickener, and the waste activated sludge is thickened with dissolved air flotation to prevent phosphorus release. The thickened sludge is combined in a sludge storage vault prior to disposal by composting. As discussed in Section 3.6, the thickened sludge liquor has been used to enhance biological phosphorus removal performance.

Turbine aerators were selected for the process reactor. Air is supplied to these turbines by centrifugal blowers. The aerators are equipped with two-speed motors so that certain cells can be operated as anoxic cells (mixing only) or aerated cells. The maximum design power consumption is about 280 kW (375 hp) for the blowers and an additional 250 kW (335 hp) for the turbine aerators, stirrers, and recycle pumps. DO probes are used to control the air supply.

The startup date for the Kelowna plant was May 1982. The total construction cost for the plant installation was about \$12,000,000 (U.S. \$). The cost reflects some unique geotechnical problems at the site, which is underlain by sedimentary and deltaic deposits.

The floors of the new structures were laid in a 3-m (10-ft) thick layer of soft silt that is contained in sand and gravel. Since a high water table was also present, extensive dewatering was required and sheet piling was used during construction (63).

The treatment performance of the Kelowna plant, described previously in Section 3.3.2, indicates the facility is able to meet its discharge requirements of 8, 7, 6 and 2 mg/l for TBOD, TSS, total nitrogen, and total phosphorus, respectively. In Section 3.6, the benefits of using primary sludge thickener liquor to enhance fermentation were described, showing that this operating procedure resulted in much lower effluent soluble phosphorus concentrations.

3.8.3 A/O Process - Pontiac, Michigan

The process design and performance of the A/O system installed at the Pontiac, Michigan East Boulevard Plant have been described in Section 3.3.3. The existing facility provided many advantages for the investigation of an A/O system in a full-scale facility. Two of four activated sludge trains were converted to the A/O process mode to allow a comparison of treatment performance with the conventional system. The A/O system was operated at cold wastewater temperatures and under nitrifying conditions, contrary to the Largo, Florida A/O operation. However, the system was also operated at a relatively long SRT for an A/O system. Waste activated sludge was anaerobically digested, and this provided an opportunity to evaluate potential phosphorus release back to the digester recycle stream. The design and performance summaries for the facility were shown previously in Tables 3-10 and 3-11, respectively.

Significant conclusions from this full-scale evaluation were that an effluent total phosphorus concentration of less than 1 mg/l could be achieved without effluent filtration and that the anaerobic digester supernatant did not present a significant recycle phosphorus load to the plant. The latter needs further investigation to explore why phosphorus was not solubilized during digestion.

It should be noted that the wastewater characteristics were very favorable for biological phosphorus removal, which tended to offset the relatively long SRT needed for nitrification. The influent TBOD:TP ratio was generally between 30 and 40.

The project also illustrated that an existing activated sludge facility could be easily upgraded for biological phosphorus removal at minimal cost and within a relatively short time period. The retrofit effort involved the allocation of 21 percent of the existing aeration basin to a three-stage anaerobic fermentation zone. This involved the addition of wooden baffles and mechanical mixers. The retrofit operation was completed in 2 months at a cost of approximately \$57,000 (1984 \$) for conversion of 13,250 m³/d (3.5 mgd) of treatment capacity to biological phosphorus removal (67). The existing aeration basin treatment capacity and air supply were sufficient to accommodate the plant modification.

3.9 Costs

The costs for biological phosphorus removal processes are sensitive to wastewater characteristics, treatment level needs, and existing equipment and site considerations. Where only phosphorus removal is required and nitrification is not occurring, reasonable retrofit treatment alternatives include chemical addition to existing biological systems and the Phostrip and A/O processes. If nitrification is

occurring or is required, the UCT process and the A/O process with an anoxic zone and internal recycle (i.e., the A²/O process) are candidates as well. An anoxic zone for partial denitrification is not strictly needed to achieve nitrification with the A/O process, but it is recommended to minimize the amount of nitrate nitrogen recycled to the anaerobic zone in the return sludge and its adverse effect on biological phosphorus release in that zone. For both phosphorus removal and a high level of nitrogen removal, the Modified Bardenpho process is a viable alternative along with a variety of advanced treatment designs using chemical addition for phosphorus removal.

An analysis was performed to compare the cost of biological phosphorus removal to that for chemical addition to activated sludge for retrofitting existing facilities (91). Effluent total phosphorus limits of 1.0 and 0.3 mg/l were considered. Nitrogen removal was not a requirement in the analysis. The analysis concluded that chemical addition to activated sludge was more cost-effective to meet a 1.0-mg/l total phosphorus effluent and was also more cost effective for meeting an effluent total phosphorus concentration of 0.3 mg/l for flows of up to 4,500 m³/d (1.2 mgd). The A/O process was determined to be more cost effective for flows of 13,600 m³/d (3.6 mgd) or more.

On the other hand, the cost of the A/O retrofit for Pontiac, Michigan, was well below these cost predictions. The Phostrip process was selected for the Reno-Sparks 150,000-m³/d (40-mgd) phosphorus removal retrofit after it was estimated that a total annual cost savings of \$500,000 would be realized compared to chemical addition to activated sludge (77). It appears, therefore, that the potential for realizing retrofit cost savings with biological phosphorus removal will likely be very site specific.

A key economic factor in the above cost analysis and in other cost analyses is the decision to include polishing filters to meet effluent phosphorus concentrations of less than 1.0 mg/l for the A/O, UCT, and Modified Bardenpho systems. This would also apply to operationally modified activated sludge processes. Some chemical addition may be required in the above processes where unfavorable BOD:P ratios exist. Cost considerations for external acetate production may also have to be developed. Previous plant performance data indicate that effluent filtration may not always be required. This will be a function of the influent BOD:P ratio or availability of fermentation products, the secondary clarifier design, the system SRT, and other parameters that affect activated sludge flocculation and clarification properties.

Cost curves for new plants have been presented in a report entitled *Emerging Technology Assessment of Biological Removal of Phosphorus* (48). Tables 3-16

through 3-19 summarize the updated capital and O&M costs developed for four basic cases. The updated capital costs are based on an Engineering News Record Construction Cost Index of 4367 (May 1987). The updated O & M costs are based on an EPA Escalation Index of 3.83. The four cases are:

Case 1:

Phosphorus removal only with a required effluent total phosphorus concentration of 1 mg/l. A comparison is made between a single-stage activated sludge system with alum addition, a Phostrip system, and an A/O system. Effluent filtration is assumed with the A/O system.

Case 2:

Same as Case 1 except the required effluent total phosphorus concentration is 2 mg/l. Without effluent filtration, in this case, the A/O system is shown to be most cost-effective.

Case 3:

Same as Case 2 with the addition of nitrification. In this case, a two-stage nitrification system with alum addition is assumed for the conventional alternative and is compared to a single-sludge A/O system. The two-stage system has a much higher capital cost.

Case 4:

Same as Case 3 with the addition of denitrification to achieve an effluent total nitrogen concentration of 3 mg/l. In this case, a three-stage activated sludge system with alum addition is compared to a Modified Bardenpho system. The three-stage system has significantly higher capital and operating costs.

In summary, the cost comparisons illustrate that the biological phosphorus removal alternatives may be competitive with conventional chemical methods. The use of effluent filtration is a critical economic factor and any final cost comparison will be extremely site specific and affected by wastewater characteristics.

Table 3-16. Cost comparison - Case 1: Phosphorus removal (effluent TP = 1 mg/l)

Alternative	Costs*	Plant Size, m ³ /d		
		1,890	18,900	189,200
1-stage AS with alum addition	Capital, \$	2,774,000	10,851,000	55,568,000
	O&M, \$/yr	218,000	868,000	5,611,000
	Total present worth, \$	4,782,000	18,846,000	107,248,000
Phostrip	Capital, \$	3,801,000	12,602,000	59,073,000
	O&M, \$/yr	273,000	744,000	3,956,000
	Total present worth, \$	6,315,000	19,455,000	95,509,000
A/O (4-hr detention) with effluent filters	Capital, \$	3,370,000	13,257,000	63,472,000
	O&M, \$/yr	227,000	836,000	4,545,000
	Total present worth, \$	5,461,000	20,957,000	105,333,000

Table 3-17. Cost comparison - Case 2: Phosphorus removal (effluent TP = 2 mg/l)

Alternative	Costs*	Plant Size, m ³ /d		
		1,890	18,900	189,200
1-stage AS with alum addition	Capital, \$	2,762,000	10,821,000	55,350,000
	O&M, \$/yr	213,000	835,000	5,276,000
	Total present worth, \$	4,724,000	18,512,000	103,944,000
Phostrip	Capital, \$	3,801,000	12,602,000	59,073,000
	O&M, \$/yr	273,000	744,000	3,956,000
	Total present worth, \$	6,315,000	19,455,000	95,509,000
A/O (4-hr detention)	Capital, \$	2,813,000	10,819,000	52,314,000
	O&M, \$/yr	197,000	692,000	3,820,000
	Total present worth, \$	4,627,000	17,193,000	87,498,000

Table 3-18. Cost comparison - Case 3: Phosphorus removal plus nitrification (effluent TP = 2 mg/l; NH₄-N = 1 mg/l)

Alternative	Costs*	Plant Size, m ³ /d		
		1,890	18,900	189,200
2-stage AS with alum addition	Capital, \$	3,370,000	12,820,000	63,381,000
	O&M, \$/yr	245,000	921,000	5,793,000
	Total present worth, \$	5,627,000	21,303,000	116,737,000
A/O (6-hr detention) for nitrification and partial denitrification to total N = 10 mg/l	Capital, \$	3,142,000	11,942,000	59,169,000
	O&M, \$/yr	210,000	764,000	4,264,000
	Total present worth, \$	5,076,000	18,979,000	98,442,000

Table 3-19. Cost comparison - Case 4: Phosphorus removal plus nitrification and denitrification (effluent TP = 2 mg/l; TN = 3 mg/l)

Alternative	Costs*	Plant Size, m ³ /d		
		1,890	18,900	189,200
3-stage AS with alum addition	Capital, \$	3,869,000	14,553,000	72,777,000
	O&M, \$/yr	296,000	1,200,000	8,059,000
	Total present worth, \$	6,595,000	25,605,000	147,004,000
Modified Bardenpho	Capital, \$	3,321,000	13,553,000	77,472,000
	O&M, \$/yr	205,000	756,000	4,552,000
	Total present worth, \$	5,209,000	20,516,000	119,398,000

* Total present worth calculated assuming a 20-year life and a discount factor of 8-7/8 percent (PWF = 9.2104)

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

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Chapter 4

Phosphorus Removal by Chemical Addition

4.1 Introduction and Theory

Many wastewater treatment plants that are required to remove phosphorus do so by adding chemicals to precipitate the phosphate present in the wastewater. As described in Section 4.2, chemicals may be added to primary, secondary, or tertiary processes, or at multiple locations in the plant. Chemicals used for phosphorus precipitation include metal salts such as ferric chloride and aluminum sulfate (alum), and lime.

This chapter: 1) describes the characteristics of chemicals used for phosphorus removal and the reactions which occur during phosphorus precipitation, 2) discusses the alternative points for chemical addition, 3) assesses the performance of chemical addition systems in removing phosphorus, and 4) provides design procedures for chemical storage and feed facilities.

A variety of metal salts are used for removal of phosphorus from municipal wastewater. The most common chemicals are aluminum sulfate (alum) and ferric chloride. Ferrous sulfate and ferrous chloride solutions, which are available as byproducts of steelmaking operations (pickle liquor) are also used. Sodium aluminate addition is sometimes practiced at facilities with low alkalinity wastewaters. Two other chemical compounds that have been investigated for phosphorus removal are aluminum chlorohydrate and polyaluminum chloride. In many cases, anionic polymers are used in addition to the mineral salt to assist in solids separation.

Mineral salts are by far the most common chemicals used for phosphorus removal. A number of plants originally designed for use of lime are currently using alum or ferric chloride. In a 1979 survey of 104 plants removing phosphorus in the lower Great Lakes basin, 53 facilities used iron salts, 49 used aluminum salts, and only 2 used lime (1).

The reasons for infrequent use of lime for phosphorus removal include: 1) the substantial increase in the mass of sludge to be handled compared to that from use of metal salts, and 2) the operation and maintenance problems associated with the handling, storage, and feeding of lime. Due to the fact that

aluminum and iron salts have all but replaced lime as a phosphorus precipitant, a detailed discussion of lime use is not presented in this Manual. Detailed information and design criteria for lime addition facilities are contained in References 2 through 5.

The reactions between phosphorus and metal salts are complex. For purposes of this discussion, it is assumed that the primary mechanism of phosphorus removal is interaction of the metal ion with orthophosphate to form an insoluble precipitate. The reactions presented below are for illustrative purposes and may not represent the true mechanisms which take place due to the variations in wastewater characteristics and forms of phosphorus present. It is recommended that any engineering examination of chemical addition for phosphorus removal include a jar test of the actual wastewater of concern (see Section 4.5.2). This will avoid the common error of assuming a required dosage when actual dosage can vary substantially between facilities (see Section 4-3) and at a given facility at different times of day and season of the year (see Section 4.5).

4.1.1 Aluminum Compounds

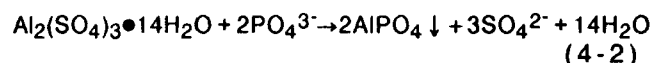
Aluminum ions combine with phosphate ions to form aluminum phosphate, as shown by:



On a mole basis, 1 mole of Al will react with 1 mole of PO_4 , or 1 mole of P. On a weight basis, 27 g of Al will react with 95 g of PO_4 (or 31 g as P) to form 122 g of AlPO_4 . The Al:P weight ratio is thus 27 Al to 31 P or 0.87:1 (2).

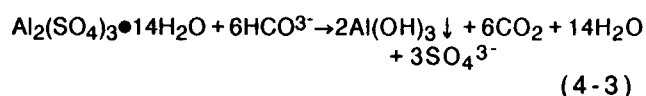
The most common form of aluminum in use for phosphorus precipitation is "alum" or "filter alum," a hydrated aluminum sulfate with the approximate formula $\text{Al}_2(\text{SO}_4)_3 \cdot 14\text{H}_2\text{O}$. Alum contains about 9.1 percent soluble aluminum as Al and 17 percent soluble aluminum as Al_2O_3 .

The reaction of alum with phosphate can be described by:



One mole (594 g) of alum will react with 2 moles (190 g) of phosphate containing 62 g phosphorus to form 2 moles (244 g) of AlPO_4 . Thus, the weight ratio of alum to phosphorus is 594 to 62 or 9.6:1.

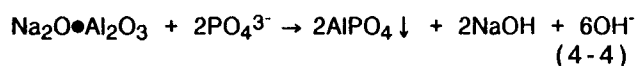
In practice, the quantities of alum required are higher than the stoichiometry would predict. This is due to competing reactions, which vary with the wastewater. Among the most notable factors that affect the actual quantity of alum required to attain a specific P concentration are: the alkalinity and final pH of the wastewater; ionic constituents such as sulfate, fluoride, sodium, etc.; quantity and nature of suspended solids, e.g., kaolin vs. montmorillonite clays; microorganisms, and other colloidal species; the actual ratio of Al to P; and the intensity of mixing and other physical conditions extant in the treatment facilities. For the purposes of engineering design the following stoichiometric reaction is provided:



The optimum pH for phosphorus removal using alum is in the range of 5.5-6.5. The extent of pH depression resulting from alum addition will depend on the alkalinity of the wastewater and the alum dosage. In unusual cases in which the buffering capacity (alkalinity) of the wastewater is very low, addition of alkaline chemicals may be required to offset the pH depression resulting from alum addition. Although strong acids could be used to lower the wastewater pH to the optimum point, it may be simpler to use a higher alum dosage to depress the pH. The relative economics of the two approaches should be evaluated by the engineer during the planning process.

Alum can be purchased as dry alum in bags, drums, or in bulk, or as liquid alum in tank cars or trucks. Characteristics of alum are given in Table 4-1 (3,4).

Sodium aluminate is sometimes used for phosphorus precipitation. The chemical formula for sodium aluminate is $\text{Na}_2\text{Al}_2\text{O}_4$ or NaAlO_2 . The commercial granular trihydrate is written as $\text{Na}_2\text{O} \bullet \text{Al}_2\text{O}_3 \bullet 3\text{H}_2\text{O}$. The reaction between sodium aluminate and phosphate may be expressed as (6):



Note the presence of NaOH as a product of the reaction, which will tend to increase pH rather than lower it. This allows sodium aluminate to be used with low alkalinity wastewaters in which use of alum would cause excessive depression of pH.

The mole ratio of Al to P is 1:1. The weight ratio of Al:P is 0.87:1, while the weight ratio of sodium

aluminate to phosphorus is approximately 3.6:1. Characteristics of sodium aluminate are given in Table 4-1 (3,4).

Aluminum chlorohydrate and polyaluminum chloride are other potentially useful chemicals for phosphorus precipitation. In jar test studies with raw wastewater samples from four treatment plants, it was found that polyaluminum chloride was superior to aluminum sulfate in removing total phosphate, while aluminum chlorohydrate gave poorer results than the other two chemicals (7). If available at a cost competitive with the more common chemicals used for phosphorus precipitation, further investigation of their use may be justified.

A "sewage grade" granular alum has been used in Scandanavia for 15-20 years and has been tested at several wastewater treatment plants in the United States. This material is a mixture of aluminum and iron sulfates, containing approximately 13.7 percent aluminum as Al_2O_3 and 4.3 percent iron as Fe_2O_3 . The wastewater treatment plant serving Geneva, New York is currently using this chemical, which was imported from Sweden. However, its distribution in the United States has been discontinued.

4.1.2 Iron Compounds

Iron salts are commonly used in the precipitation of phosphorus from municipal wastewater. Both ferrous (Fe^{2+}) and ferric (Fe^{3+}) ions can be used in the form of ferric chloride, ferrous chloride, ferric sulfate, and ferrous sulfate. Ferrous chloride and ferrous sulfate are also available as byproducts of steelmaking operations (waste pickle liquor), although these solutions may contain large quantities of free hydrochloric or sulfuric acid which can cause destruction of alkalinity and pH depression. Characteristics of iron salts used for phosphorus precipitation are given in Table 4-1 (3,4).

A typical reaction between ferric chloride and phosphate can be approximated by (8):



The mole ratio of Fe to P is 1:1. 162.3 g of FeCl_3 will react with 95 g of PO_4 to form 150.8 g of FePO_4 . Stoichiometric weight ratio of Fe:P is 1.8:1, while the weight ratio of FeCl_3 to P is 5.2:1. As with alum, the reaction mechanism is more complex than the equation shown above.

The reaction between ferrous salts (ferrous chloride and ferrous sulfate) and phosphate can be approximated by:

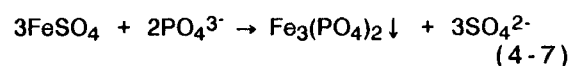
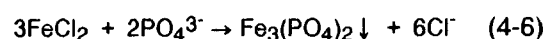
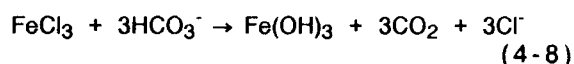


Table 4-1. Characteristics of Aluminum and Iron Salts (3,4)

Common Name and Formula	Shipping Data		Physical and Chemical Characteristics		
	Available Forms	Containers and Requirements	Appearance and Properties	Bulk Density (kg/cm)	Commercial Strength
Dry alum $\text{Al}_2(\text{SO}_4)_3 \cdot 14\text{H}_2\text{O}$	Lump Ground Rice Powdered	Bags: 45,90 kg; Bbl.: 135,180 kg; Drums: 11,45,110 kg; Bulk - Car loads Tank trucks Store dry	White/cream color. pH: 3.0-3.5 for 1-10% solution. Dust is irritant to mucous membranes.	600-1,200	17% Al_2O_3 by wt.
Liquid alum $\text{Al}_2(\text{SO}_4)_3 \cdot 14\text{H}_2\text{O}$		Bulk - Car loads Tank trucks	Will begin to crystalize @ -1°C ; Crystallizes @ -8°C . Corrosive.	1,330 @ 16°C	8.3% Al_2O_3 by wt.
Dry sodium aluminate $\text{Na}_2\text{Al}_2\text{O}_4$		Bags: 23,45,68 kg; Bulk: not available. 6 month max. storage.	pH: 11.9 for 1% solution. Non-corrosive. Dust is irritant.	640-800	41-46% Al_2O_3 by wt.
Liq. sodium aluminate $\text{Na}_2\text{Al}_2\text{O}_4$		Drums: 170 kg; Tank truck; Tank car; 2-3 month max. storage.	Strong alkali. Handle as caustic.		4.9-26.7% Al_2O_3 by wt.
Liq. ferric chloride FeCl_3		Carboys: 19,49 l; Tank trucks: 11,500-15,000 l; Tank cars: 15,000-38,000 l	Dark brown, oily. pH: 2.0 for 1% solution. Very corrosive., stains concrete and other materials.	1,340-1,490	35-45% FeCl_3 by wt.
Liq. ferrous chloride FeCl_2	Commercial; waste pickle liquor	Drums: 190 l; Tank trucks: 15,000 l Tank cars	Dark brown, oily. Free acid content typically 1-1.5% but may reach 10%. Slightly less corrosive than FeCl_3	1,190-1,250	20-25% FeCl_2 by wt.
Dry ferrous sulfate $\text{FeSO}_4 \cdot 7\text{H}_2\text{O}$	Commercial; waste pickle liquor		Acidic when dissolved. Composition is variable. Oxidizes in moist air. Cakes @ storage temp. above 20°C .	990-1,060	55-58% FeSO_4 by wt.

In reality these reactions are more complex. The mole ratio of Fe to P is 3:2. Weight ratio of ferrous ion to phosphorus is 3.2:1.

Addition of iron salts will result in the destruction of alkalinity as described by:



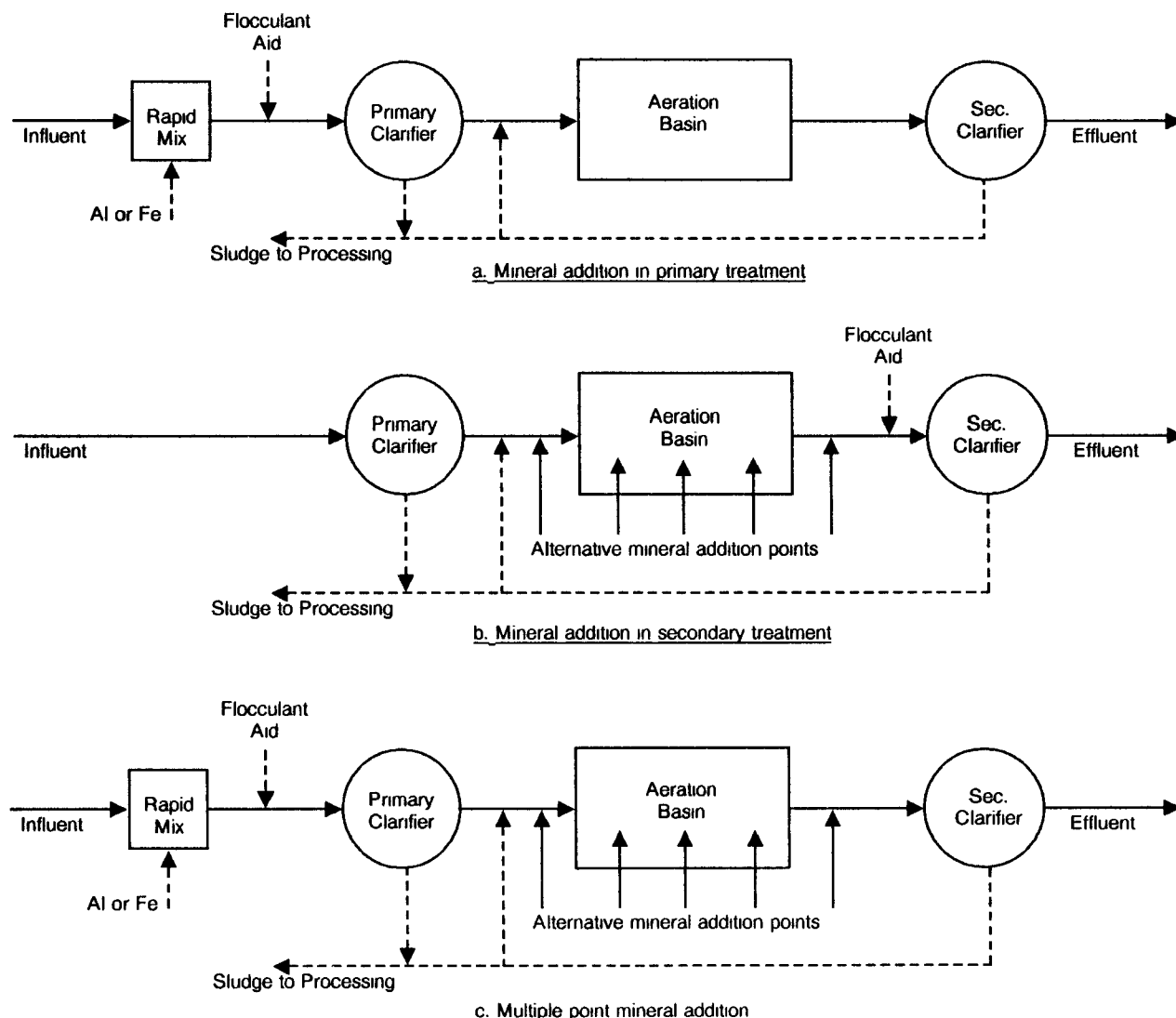
Iron salts are most effective for phosphorus precipitation within a certain pH range. For ferric (Fe^{3+}) ion, the optimum pH range is 4.5-5.0. However, significant removal of phosphorus can be achieved at higher pH. For ferrous (Fe^{2+}) ion, the optimum pH is approximately 8. Good phosphorus removal can be achieved between pH values of 7 and 8. Canadian studies have shown that effective phosphorus precipitation does not occur until the ferrous ion is oxidized to ferric ion, and for this reason do not support the use of ferrous salts in primary treatment (9). However, both ferrous chloride and ferrous sulfate have been used effectively in primary treatment (3,10).

4.2 Application Points

The most common points for addition of aluminum and iron salts for phosphorus removal are: 1) immediately upstream of the primary clarifier, 2) in or immediately after the aeration basins prior to final clarification, and 3) at both points simultaneously. Another option is use of separate, tertiary chemical clarification. Such a scheme would only be justified for very stringent effluent discharge standards, as for reuse.

Figure 4-1 shows the common schemes for chemical addition in an activated sludge plant. The advantages of addition to primary treatment include greater opportunity for adequate mixing and flocculation, and reduced loadings to downstream processes as a result of improved BOD and SS removal. The major disadvantage of chemical addition to the primaries is that incomplete phosphorus precipitation may result because of the presence of phosphorus forms other than orthophosphate that are not easily precipitated.

Figure 4-1. Alternative schemes for mineral addition for phosphorus removal.



In general, higher levels of phosphorus removal can be achieved by chemical addition to the secondary process or by addition at multiple points in the treatment train.

As discussed in Chapter 2, adequate wastewater characterization and jar testing are essential before implementing a chemical addition program.

Table 4-2, compiled based on a survey of 104 plants in the U.S. and Canada, shows a breakdown of plants by location of chemical addition and by cation used for precipitation (1).

4.2.1 Mineral Addition Before Primary Clarification

Phosphorus removal by addition of mineral salts in primary treatment requires good mixing and flocculation in order to ensure optimum results. With

proper design and operation of mixing and flocculation systems, 70-90 percent phosphorus removal can be achieved in primary treatment. Significant increases in BOD and SS removal efficiencies can also be expected from mineral salt addition. Table 4-3 provides a summary of potential removal efficiencies of P, BOD, and SS in primary and secondary treatment with and without mineral addition (3).

For mineral addition to primary treatment, provision of a separate rapid mix tank may be necessary. As discussed in Section 4.2.4, existing hydraulic structures such as Parshall flumes and drop manholes have been used, but may not provide sufficient turbulence to ensure complete and intimate mixing of the chemical with the wastewater. Good performance has been reported in Canada by injection into the discharge side of raw sewage pumps, addition to aerated grit chambers, preaeration

Table 4-2. Distribution of Selected Phosphorus Removal Facilities by Point of Chemical Addition and Cation Used (1)

Point of Addition	Number of Plants								
	United States			Canada			Total		
	Al	Fe	Total	Al	Fe	Total	Al	Fe	Total
Primary	1	16	17	2	20	22	3	36	39
Secondary	26	6	32	17	8	25	43	14	57
Tertiary	2	2	4	1	0	1	3	2	5
Total	29	24	53	20	28	48	49	53 ¹	102

¹ One plant does not specify point of addition.

Table 4-3. Potential effectiveness of Primary and Secondary Treatment With and Without Mineral Addition for Phosphorus Removal

	Phosphorus Removal (%)		SS Removal (%)		BOD Removal (%)	
	Without	With	Without	With	Without	With
Primary Treatment	5-10	70-90	40-70	60-75	25-40	40-65
Secondary Treatment						
Trickling Filter	10-20	80-95	80-90	85-95	75-90	80-95
Activated Sludge	10-20	80-95	80-95	85-95	85-95	85-95

channels, or Parshall flumes, and addition to pipes or channels between primary clarifiers and aeration tanks if supplemented with mechanical or air mixing (11,12).

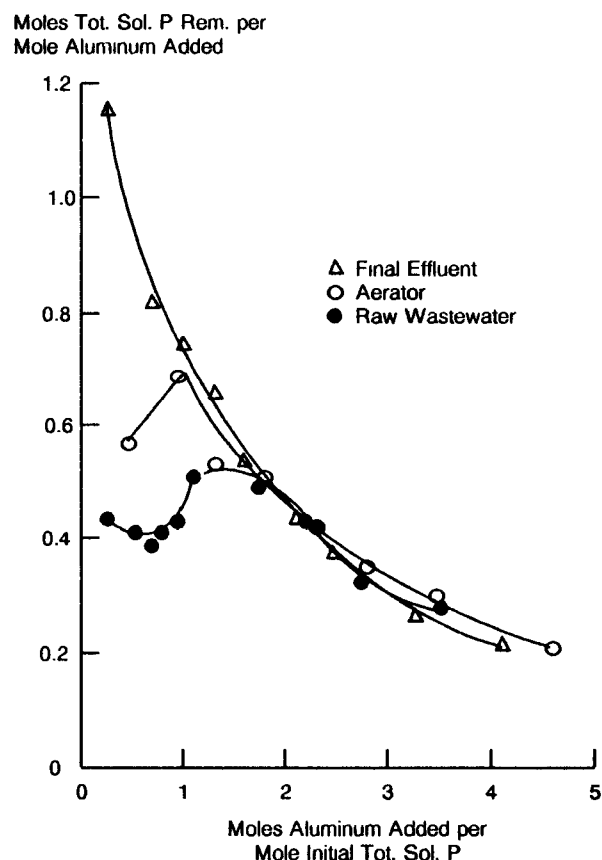
Initial mixing intensity is very important for complete dispersal of the coagulant and, thus, maximum efficiency. Flocculation normally occurs naturally in the primary clarifier, particularly in the center feed well, which may preclude the need for a separate flocculation basin. New clarifiers can be designed with mixers and a center flocculation well with as much as 30 minutes detention time. If necessary, existing clarifiers can be modified to provide a designated flocculation zone.

Due to competing reactions and variation in coagulant demand, more chemical will generally be required for phosphorus precipitation in primary treatment than in secondary or tertiary treatment. This can be seen in Figure 4-2, which shows phosphorus removed per mole of aluminum added for three different application points (3).

4.2.2 Mineral Addition to Secondary Processes

Addition of aluminum or iron salts directly to the aeration basins or between the aeration basins and final clarifiers is common practice for chemical precipitation of phosphorus. This alternative has considerable flexibility in the point of chemical addition, allowing modifications of the injection point to ensure use of the best available conditions for coagulation and flocculation to occur. Unfortunately, the approach has some drawbacks, in that velocity

Figure 4-2. Impact of point of addition on effectiveness of phosphorus removal using aluminum (3).



gradients or turbulence levels are likely to be less than ideal for proper mixing and flocculation to occur. Since the optimal point of chemical addition will vary depending on the choice of chemicals, velocity gradients in the aeration basin and inter-basin channels, and wastewater characteristics, full scale experimentation with various points of addition will likely be necessary. If a high degree of phosphorus removal is required, mineral addition should occur downstream of any return streams such as digester supernatant. Addition of minerals to secondary processes may result in an increase in dissolved solids in the effluent, particularly when pickling liquors or other "impure" chemical sources are used.

4.2.3 Mineral Addition at Multiple Points

Addition of mineral salts at multiple locations in the treatment plant has been found to be an efficient and cost-effective means of chemical addition for phosphorus control. Advantages of this approach are overall reduction in chemical requirements to achieve a given effluent phosphorus objective and increased operational flexibility. In design of new facilities, provision of multiple chemical addition points is recommended to allow optimization of the chemical feed system to achieve the most economical and reliable solution.

It should be noted that, when aluminum or iron salts are used for phosphorus precipitation, addition of small amounts of a coagulant aid such as anionic polyelectrolytes may be necessary before the final clarifier to assist in removing dispersed metal-phosphate floc. A typical polymer dose when used as a coagulant aid is 0.1-0.25 mg/l (8).

4.3 Performance

Table 4-4 shows performance data from selected facilities using various mineral salts and points of addition to achieve phosphorus removal (13). In general, 70-90 percent phosphorus removal can be expected in primary treatment, with removal efficiencies in secondary treatment ranging from 80 to 95 percent.

Dosages can be expected to vary widely depending on choice of chemical and wastewater characteristics, as well as other factors such as degree of mixing intensity at the point of addition and opportunity for flocculation. Table 4-5 shows a summary of dosages employed during full-scale studies at numerous wastewater treatment plants (14). The dosages cited are those required to achieve a total phosphorus concentration of 1 mg/l in the plant effluent.

4.4 Equipment Requirements

4.4.1 Chemical Handling and Storage

As seen in Table 4-1, a variety of chemicals are available for precipitation of phosphorus from

wastewater. Each chemical has characteristic requirements regarding type and maximum duration of storage, choice of piping and transport systems, chemical feeding equipment, and safety precautions. Chemical handling and storage requirements for common chemicals used in phosphorus control are described below.

a. Aluminum Salts

Aluminum sulfate is available in dry or liquid form. Dry alum or "filter alum" can be purchased in lump, ground, rice, or powdered grades. Water utilities prefer ground or rice alum because of superior flow characteristics. Bulk alum should be stored in mild steel or concrete bins with dust collection equipment. Gates should be provided on bulk storage vessels to allow isolation of feed equipment.

A typical bulk storage tank for dry chemicals is shown in Figure 4-3. Bulk dry alum can be transferred with screw conveyors, bucket elevators, or pneumatic conveyors. Bags and drums of alum should be stored in dry locations. Day hoppers receiving alum from bags or drums should have a minimum storage capacity of eight hours at the maximum expected feed rate. Hopper bottoms should have a minimum wall slope of 60 degrees to prevent arching.

Dry alum is not corrosive unless it absorbs moisture. Alum dust, however, can cause minor irritation of the eyes and respiratory tract.

Liquid alum is available in 11-19 m³ (3,000-5,000 gal) tank truck lots or 26-68 m³ (7,000-18,000 gal) tank car lots. Transportation costs are greater for liquid alum since it is nearly half water by weight. Liquid alum will generally be more economical than dry alum if the point of use is within 160 km (100 mi) of the manufacturing location. However, because of the ease of handling, storage, and feeding in liquid form, the practical limit for transport may be 320 km (200 mi) or more (3).

Alum is typically stored without dilution at the shipping concentration received at the plant. Storage tanks located outside should be closed and vented, with provisions for heating to maintain temperatures above -4°C (25°F) to prevent crystallization. Materials of construction for liquid alum storage vessels include type 316 stainless steel, fiberglass-reinforced plastic (FRP), or steel lined with rubber, polyvinyl chloride (PVC), or lead. Liquid alum can be stored indefinitely. Storage tanks should be sized to accommodate a 10-day to 2-week supply and should be capable of handling 1-1/2 times the maximum quantity shipped.

Liquid alum is moderately corrosive, and hand and face protection should be worn when working on leaking equipment. Any spills should be immediately flushed with water as liquid alum becomes very slippery upon evaporation (3).

Table 4-4. Performance of Facilities Using Mineral Salts for Phosphorus Removal (11)

Plant Type and Location	Design Flow	Average Flow	Chemicals	Chemical Feed Point	Chemical Dosage	Metal Ion: Inf. TP	Inf. TP	Eff. TP
	m ³ /d	m ³ /d			mg/l as Metal Ion		mg/l	mg/l
<u>Plug Flow AS</u>								
Waupaca, WI	4,760	2,200	Alum	Sec. Clarifier	24.6	3.25	7.56	0.86
East Chicago, IN	75,700	59,800	Alum	Sec. Clarifier	7.7	3.99	1.93	0.38
			Polymer	Sec. Clarifier	1.0			
Mason, MI	5,700	5,000	Ferric Chloride	Prim. Clarifier	9.1	1.4	6.5	0.88
			Polymer	Prim. Clarifier	0.05			
Flushing, MI	4,400	6,000	Ferric Chloride	Sec. Biol. Process	5.3	1.56	3.4	0.48
				Sec. Biol. Process	0.15			
Appleton, WI	62,500	52,200	Ferrous Chloride	Plant Influent	16.8	1.6	10.5	0.8
Grand Ledge, MI	5,700	3,000	Ferrous Chloride	Sec. Biol. Process	5.6	1.24	4.5	0.7
Bowling Green, OH	30,300	20,100	Ferrous Chloride	Sec. Clarifier	5.2	0.62	8.4	0.75
			Polymer	Sec. Clarifier				
Kenosha, WI	106,000	90,500	Ferrous Sulfate	Prim. Clarifier	5.35	1.43	3.74	0.36
Toledo, OH	386,100	310,400	Ferrous Sulfate	Prim. Clarifier	3.6	1.3	2.76	0.35
			Polymer	Prim. Clarifier				
Clintonville, WI	3,800	2,700	Ferrous Sulfate	Sec. Clarifier	5.3	1.47	3.6	0.75
<u>Complete Mix AS</u>								
Thiensville, WI	900	3,300	Alum	Sec. Biol. Process	9.3	2.46	3.78	0.29
			Polymer	Sec. Biol. Process	0.82			
Two Harbors, MN	4,500	3,400	Alum	Sec. Clarifier	9.6	1.6	6.0	0.25
Escanaba, MI	8,300	7,600	Ferric Chloride	Prim. Clarifier	4.7	1.04	4.5	0.82
			Polymer	Prim. Clarifier	0.35			
Sheboygan, WI	69,600	46,600	Ferric Chloride	Sec. Clarifier	10.2	1.6	6.38	0.9
Lima, OH	70,000	15,100	Ferrous Chloride	Prim. Clarifier	13.2	3.38	3.9	0.5
			Polymer	Prim. Clarifier	0.07			
Niles, MI	22,000	12,100	Ferrous Chloride	Sec. Biol. Process	10.9	2.66	4.1	0.7
Crown Point, IN	13,600	8,700	Ferrous Chloride	Sec. Clarifier	11.0	2.0	5.5	0.7
			Polymer	Sec. Clarifier	0.94			
Cedarburg, WI	11,400	7,600	Ferrous Sulfate	Sec. Clarifier	9.9	2.99	3.31	0.67
			Polymer	Sec. Clarifier				
<u>Contact Stabilization AS</u>								
Neenah, WI	5,700	4,000	Alum	Prim. Clarifier	7.7	2.2	3.5	0.7
Neenah, WI	14,800	16,700	Alum	Sec. Biol. Process	4.1	1.0	4.1	0.8
Algoma, WI	2,800	3,000	Ferric Chloride	Prim. Clarifier	33.0	10.0	3.3	0.23
			Polymer	Prim. Clarifier	0.07			
Grafton, WI	8,100	3,600	Ferrous Chloride	Prim. Clarifier	16.2	2.31	7.0	0.69
Port Washington, WI	4,700	5,800	Ferrous Chloride	Prim. Clarifier	8.5	1.44	5.9	1.0
Port Clinton, OH	5,700	6,400	Ferrous Chloride	Sec. Biol. Process	10.2	1.96	5.2	0.5
Oberlin, OH	5,700	5,700	Ferrous Chloride	Prim. Clarifier	6.4	1.08	5.9	1.0
North Olmstead, OH	34,000	21,200	Sodium Aluminate	Sec. Biol. Process	8.3	2.86	2.9	0.7
<u>Pure Oxygen AS</u>								
Fon du Lac, WI	41,600	26,900	Alum	Sec. Clarifier	8.5	1.18	7.2	0.73
			Polymer	Sec. Clarifier	0.75			
<u>Extended Aeration AS</u>								
Aurora, MN	1,900	1,700	Alum	Prim. Clarifier	16.9	5.83	2.9	0.76
Upper Allen, PA	1,800	1,200	Alum	Sec. Biol. Process	8.2	0.92	8.9	2.0
			Polymer	Sec. Biol. Process	0.37			
Corunna, Ontario	3,800	2,000	Alum	Sec. Clarifier	5.0	0.65	7.74	0.36
Saukville, WI	7,600	2,400	Ferrous Chloride	Prim. Clarifier	10.3	1.61	6.4	0.59
Plymouth, WI	6,200	5,800	Ferrous Chloride	Sec. Biol. Process	7.7	1.15	6.7	0.77
Trenton, OH	13,200	9,600	Ferrous Chloride	Sec. Biol. Process	2.56	0.42	6.1	0.65
Seneca, MD	18,900	15,100	Sodium Aluminate	Plant Influent	4.3	0.61	7.1	1.6
			Polymer	Sec. Clarifier	2.4			

Table 4-4. Performance of Facilities Using Mineral Salts for Phosphorus Removal (continued)

Plant Type and Location	Design Flow	Average Flow	Chemicals	Chemical Feed Point	Chemical Dosage	Metal Ion: Inf. TP	Inf. TP	Eff. TP
	m ³ /d	m ³ /d			mg/l as Metal Ion		mg/l	mg/l
<u>Step Aeration AS</u>								
Fort Wayne, IN	227,100	170,100	Ferrous Chloride	Sec. Biol. Process	4.3	0.54	7.9	0.67
East Lansing, MI	71,200	42,800	Ferrous Chloride Polymer	Sec. Clarifier Sec. Clarifier	5.9 0.05	1.11	5.3	0.9
Oak Creek, WI	454,200	340,650	Ferrous Sulfate	Sec. Biol. Process	4.4	0.96	4.6	0.54
Elkhart, IN	75,700	60,200	Ferrous Sulfate	Sec. Clarifier	1.6	0.63	2.56	0.83
<u>2-Stage Nitrification AS</u>								
Piscataway, MD	113,600	54,900	Alum Polymer	Sec. Clarifier Sec. Clarifier	8.8 3.8	1.44	6.13	0.2
<u>High Rate TF</u>								
Geneva, OH	7,600	3,900	Alum	Sec. Clarifier	12.1	4.03	3.0	0.4
Coldwater, MI	8,700	7,400	Ferric Chloride Polymer	Sec. Clarifier Sec. Clarifier	8.3 0.1	2.02	4.1	0.88
Oconto Falls, WI	1,900	1,400	Ferric Chloride	Sec. Biol. Process	8.81	2.4	3.67	0.45
Kendalville, IN	10,100	7,600	Ferric Chloride Polymer	Sec. Biol. Process Sec. Biol. Process	14.7 0.25	4.05	3.63	0.35
<u>Standard Rate TF</u>								
Willard, OH	5,100	4,800	Alum Polymer	Prim. Clarifier Prim. Clarifier	6.3 0.14	1.21	5.2	0.82
Elizabethtown, PA	11,400	6,500	Alum Polymer	Sec. Clarifier Sec. Clarifier	12.8 0.4	2.51	5.1	1.7
Durand, MI	3,000	2,700	Ferric Chloride	Prim. Clarifier	11.2	2.2	5.1	0.83
Saginaw, MI	16,700	6,400	Ferric Chloride Polymer	Prim. Clarifier Prim. Clarifier	9.6 0.1	0.99	9.7	1.5
Little Hunting Creek, VA	17,000	14,400	Ferric Chloride Polymer	Prim. Clarifier Prim. Clarifier	42.5 2.8	4.57	9.3	0.2
Bay City, MI	75,700	33,300	Ferric Chloride Polymer	Sec. Clarifier Sec. Clarifier	9.5 0.29	2.07	4.6	0.5
Coloma, MI	8,300	5,300	Ferrous Chloride	Prim. Clarifier	4.1	1.71	2.4	0.65
<u>RBC</u>								
Romeo, MI	6,100	3,300	Alum Polymer	Prim. Clarifier Prim. Clarifier	7.1 0.77	2.4	2.96	0.46
Chesaning, MI	2,200	2,000	Ferric Chloride Polymer	Prim. Clarifier Prim. Clarifier	9.0 0.4	3.46	2.6	0.6
Negaunee, MI	6,100	3,300	Ferric Chloride Polymer	Prim. Clarifier Prim. Clarifier	7.5 1.0	3.75	2.0	0.95
Dexter, MI	2,200	800	Ferric Chloride Polymer	Sec. Clarifier Sec. Clarifier	10.2 0.5	2.0	5.11	0.46
Hartford, MI	1,300	800	Ferrous Chloride Polymer	Prim. Clarifier Prim. Clarifier	13.0 0.6	3.25	4.0	0.75
St. Johns, MI	7,200	6,300	Ferrous Chloride Polymer	Prim. Clarifier Prim. Clarifier	5.01 0.04	1.38	3.7	0.5
Charlotte, MI	4,500	2,700	Ferrous Chloride Polymer	Prim. Clarifier Sec. Clarifier	13.7 0.18	2.45	5.6	0.68
<u>Oxidation Ditch</u>								
Lapeer, MI	7,000	7,200	Ferric Chloride	Sec. Clarifier	4.65	0.88	5.3	1.2
Portage, IN	13,200	8,400	Ferrous Chloride	Sec. Clarifier	9.9	1.65	6.0	1.5

Table 4-5. Chemical Dosage Summary from Ontario Treatability Studies (14)

Point of Addition	Chemical	Number of Plants	Ave. Dosage ^{a,b}	Ave. Metal Ion/TP Ratio
			mg/l	
Raw wastewater	Ferric chloride	7	14.2	2.7
	Alum	5	10.3	1.7
Mixed liquor	Ferric chloride	20	9.5	1.5
	Alum	15	7.5	1.6

^a Dosage required to achieve effluent TP of 1 mg/l.

^b Expressed as Fe or Al.

Sodium aluminate is available in both dry and liquid forms (see Table 4-1). Dry sodium aluminate should be stored for a maximum of six months at 16-32°C (60-80°F) and will deteriorate with exposure to the atmosphere. Hopper agitation may be required to prevent caking and bridging. Storage vessels may be mild or stainless steel, FRP, or concrete. Use of copper and its alloys, rubber, and aluminum should be avoided. Maximum recommended storage time for liquid sodium aluminate is two to three months (4).

Sodium aluminate should be treated as a caustic similar to sodium hydroxide. Body contact should be prevented, and hand and face protection must be worn when working on sodium aluminate storage or feed equipment.

b. Iron Salts

Iron salts discussed here include ferric chloride, ferrous chloride, and ferrous sulfate, which are the most common iron compounds used for phosphorus removal. Other iron salts such as ferric sulfate can also be employed for precipitation of phosphorus.

Ferric chloride is available as a liquid in carboys and in bulk, as shown in Table 4-1. Tank trucks and tank cars are usually unloaded pneumatically. Designated safety procedures should be closely followed during loading and unloading operations. Ferric chloride storage tanks may be constructed of steel lined with rubber or plastic, FRP, or synthetic resins. In most cases, ferric chloride should be stored in heated buildings or in heated tanks to prevent crystallization. Liquid ferric chloride can be stored indefinitely without deterioration. Ten to 14 days of storage capacity is recommended, with the ability to handle 1-1/2 times the largest anticipated shipment (3).

Ferric chloride is a corrosive material. When working on ferric chloride handling equipment, workmen should wear rubber gloves, rubber aprons, and goggles or a face shield. Any contact with eyes or skin should be flushed with running water. If ingested, vomiting should be induced.

Ferric chloride will stain concrete and other materials. To prevent staining in areas where ferric chloride is handled, rubber mats or resistant coatings should be used.

Ferrous chloride, or waste pickle liquor, is a byproduct of steelmaking operations and is available in bulk tank car or tank truck lots. The free acid content may vary from 1 to 10 percent but is usually 1-2 percent. Although slightly less corrosive than ferric chloride, ferrous chloride generally has the same storage and handling requirements. Since pickle liquor may not be available on a continuous basis, storage and handling facilities should be suitable for accommodating ferric chloride as an alternate chemical.

Ferrous sulfate is also a byproduct of steelmaking operations, although the product is normally sold in the dry form as granules, crystals, powder, or lumps. Composition may be variable but typically contains 55-58 percent FeSO₄. Dry ferrous sulfate will oxidize and hydrate in moist air and will cake at temperatures above 20°C (68°F). In the dry form, it should be stored in cool dry areas. Storage containers may be constructed of concrete, synthetic resin, or steel lined with asphalt, rubber, PVC, or chemically resistant resins (3,4). Ferrous sulfate dust is irritating to the eyes and respiratory tract.

Ferrous sulfate solution is acidic and should be handled with the same precautions that apply to ferric chloride. Construction materials for storage vessels for ferrous sulfate solutions are the same as those for ferric chloride solutions.

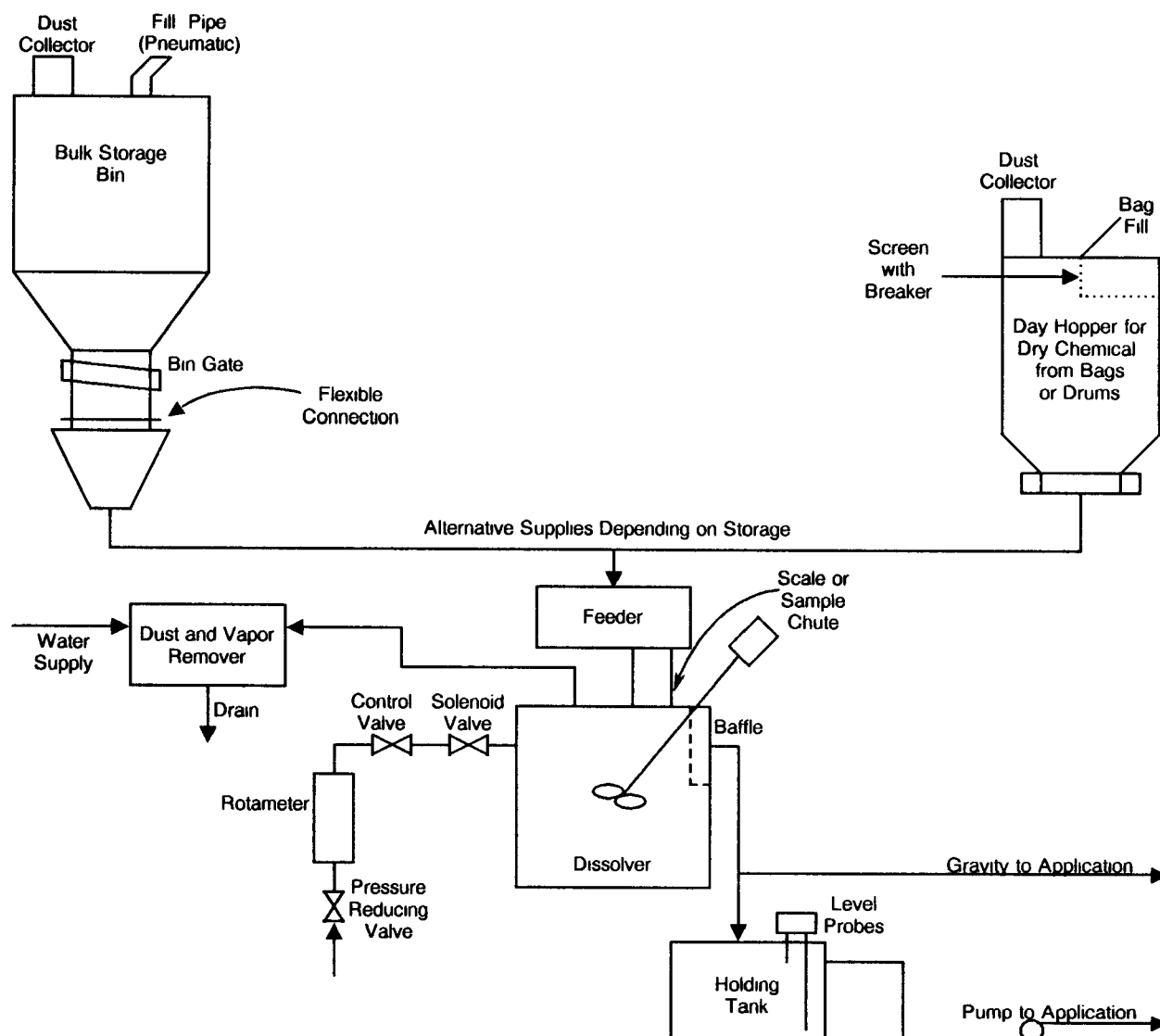
c. Polymer

While polymer is not used specifically for phosphorus precipitation, its use in conjunction with aluminum and iron salts is so common as to warrant a separate discussion. Small quantities of polymer are typically added just downstream of the metal addition point to assist in the agglomeration and settling of the metal-phosphate floc. A minimum lag time of 10 seconds (flow time) is recommended between the point of metal addition and the polymer injection point (14). Some designers have suggested lag times of two to five minutes (15).

A multitude of polymers are available in dry or liquid form. Dry polymer is shipped in a variety of packages and containers, depending on the manufacturer. Bagged dry polymer should be stored in cool, low humidity areas. Bags should be removed in proper rotation to prevent excessive storage times. Polymers are generally low in toxicity and are not irritating.

Polymer is added to a process in the solution form, requiring blending of dry polymer with water to form a stock solution, followed by an aging period. The stock solution is usually diluted prior to use. Liquid polymers

Figure 4-3. Typical dry chemical feed system.



require no aging but are normally diluted with water before application.

Polymer solutions are typically stored in FRP, type 316 stainless steel, or plastic lined steel tanks. To prevent deterioration, polymer solutions should be stored no longer than 1-3 days. Check with the manufacturer for recommendations for the specific product.

Instantaneous blending polymer systems may have application for temporary or intermittent use situations. These systems automatically meter, dilute, activate, and feed liquid polymer and water and do not require separate storage, holding, and mixing tanks.

4.4.2 Dry Chemical Feeding and Dissolving

Dry chemical feed equipment can be of three types:

- volumetric
- loss-in-weight gravimetric
- belt gravimetric

Volumetric feeders are the least expensive and can be used where cost is a concern, chemical delivery rates are low, and great accuracy is not required. Volumetric feeders generally employ a screw feed mechanism.

Loss-in-weight gravimetric feeders provide a high degree of accuracy (1 percent) and are recommended where close control of chemical

dosages can result in substantial savings in chemical costs. Maximum feed rates of such units are approximately 1,800 kg (4,000 lb)/hr.

Belt gravimetric feeders are intermediate in cost between volumetric and loss-in-weight gravimetric feeders, and can provide accurate and reliable service.

In general, closed construction is preferable for chemical feeders, since this exposes a minimum of operating components to the corrosive vapors from the dissolving or solution tank. The various types of chemical feeders available are shown in Table 4-6 (4).

Gravimetric feeders offer the following advantages over volumetric feeders:

1. Calibration is normally not required.
2. Greater accuracy and dependability.
3. Incorporation of totalizer to allow maintenance of accurate records and inventories.
4. Automatic proportioning.
5. Low maintenance; simple operation.

When dry chemicals are used, a working solution is made up by blending water with the chemical in a mechanically-agitated dissolving tank or solution tank. With bulk chemicals, such systems employ a water meter in conjunction with a variable rate feeder to achieve a continuous stream of the solution at the proper strength. With bags or containers, the proper solution is made up manually on a batch basis.

With alum or sodium aluminate, the recommended minimum solution strength is 6 percent or 0.06 kg/liter of water (0.5 lb/gal). The detention time in the dissolver should be 5 minutes at the maximum feed rate. The same recommendations are made for dissolution of dry ferrous sulfate. For ferric sulfate, feed solutions are made up at a water to chemical weight ratio of 2:1 to 8:1, with a typical ratio being 4:1 or 0.25 kg Fe_2SO_4 /liter of water (2.1 lb/gal). Solutions having strengths less than 1 percent are subject to hydrolysis and deposition of ferric hydroxide.

A typical arrangement for the feeding and dissolving of dry chemicals is shown in Figure 4-3. It should be noted that the degree of automation in dry chemical dissolution systems will depend on the size of the plant and daily chemical usage.

For plants less than 3,785 m³/d (1 mgd), manual preparation of the chemical solution on a batch basis may be indicated. This is typically accomplished in a day tank in which dry, bagged chemical is mechanically mixed with water to reach the desired concentration. For larger facilities, the chemical solution is prepared automatically using a controller which adjusts feed rate of dry chemical in proportion

to potable water flowrate (16). A system of this type is depicted in Figure 4-3.

4.4.3 Liquid Chemical and Solution Feeding

Several alternatives are available for feeding liquid chemicals or chemical solutions. The pressure head often determines the type of system to be used. Rotary dipper feeders or rotameters with control valves are commonly used for gravity feed applications, while metering pumps are used for feeding chemical solutions under pressure.

Although provision is sometimes made for dilution of liquid alum or ferric chloride prior to feeding into the process, this is generally unnecessary, and may be undesirable due to the occurrence of hydrolysis at dilute concentrations. It has been found that feeding undiluted liquid alum results in better coagulation and settling (3). For polymers, dilution of the stock solution is generally practiced to allow better dispersion of the polymer in the wastewater.

Figure 4-4 shows typical chemical feed arrangements for elevated chemical storage systems, while Figure 4-5 shows alternatives for ground storage. Rotary dipper feeders are reliable feeders that are commonly used for gravity flow applications. Feed rates can be varied based on a signal from a mainstream flow meter (flow proportional control) as discussed later in this section. Rotameters in conjunction with control valves may also be used for small applications where frequent variation in chemical feed rate is not required. Rotameters should not be used with ferric chloride or other iron solutions since the sight glass will become stained and opaque.

Centrifugal transfer pumps, as shown in Figure 4-5a, should be direct connected but not close-coupled to prevent leakage into the motor. Pump components for liquid alum service should be constructed of 316 stainless steel, FRP, or plastics. For ferric chloride, graphite or rubber lined pumps with Teflon seals are recommended. Metering pumps are typically of the positive displacement type, either diaphragm or plunger. Diaphragm pumps protected with internal or external relief valves are preferred. A back pressure valve is recommended to provide positive check valve operation (3). Materials of construction for chemical feed service include 316 stainless steel, FRP, plastics, and rubber. Manufacturers recommendations should be followed regarding selection of pump materials for the specific chemical of interest.

For pipes transporting alum solution, use of FRP, PVC, or other plastics is recommended. Valves should be plastic, 316 stainless steel, or rubber-lined iron or steel. For ferric chloride conveyance, pipes should be constructed of steel lined with rubber or Saran, FRP, or plastics. Valves should be rubber- or resin-lined diaphragm valves, Saran-lined valves with Teflon diaphragms, rubber-sleeved pinch

Table 4-6. Types of Chemical Feeders (4)

Type of Feeder	Use	Limitations	
		Capacity liters/h	Range
Dry feeders			
Volumetric:			
Oscillating plate	Any material, granules or powder.	0.3 - 1,000	40 to 1
Oscillating throat (universal)	Any material, any particle size.	0.06 - 2,800	40 to 1
Rotating disc	Most materials including NaF, granules or powder.	0.3 - 28	20 to 1
Rotating cylinder (star)	Any material, granules or powder.	230 - 57,000 or 200 - 8,500	10 to 1 or 100 to 1
Screw	Dry, free flowing material, powder or granular.	1.4 - 510	20 to 1
Ribbon	Dry, free flowing material, powder, granular or lumps.	0.06 - 4.5	10 to 1
Belt	Dry, free flowing material up to 1.5-in size, powder or granular.	2.8 - 85,000	10 to 1 to 100 to 1
Gravimetric:			
Continuous - belt and scale	Dry, free flowing granular material, or floodable material.	0.6 - 57	100 to 1
Loss in weight	Most materials, powder, granular or lumps.	0.6 - 2,300	100 to 1
Solution feeders			
Nonpositive displacement:			
Decanter (lowering pipe)	Most solutions or light slurries.	0.3 - 280	100 to 1
Orifice	Most solutions.	4.5 - 142	10 to 1
Rotameter (calibrated value)	Clear solutions.	0.1 - 4.5 or 0.3 - 570	10 to 1
Loss in weight (tank w/control valve)	Most solutions.	0.06 - 5.7	30 to 1
Positive displacement:			
Rotating dipper	Most solutions or slurries.	2.8 - 850	100 to 1
Proportional pump:			
Diaphragm	Most solutions. Special unit for 5% slurries. ¹	0.1 - 4.2	100 to 1
Piston	Most solutions, light slurries.	0.3 - 4,800	20 to 1

¹ Use special heads and valves for slurries.

valves, or plastic ball valves (3,4). Pipe selection for polymer service should be made after the type of polymer has been determined. Plastic pipe or type 316 stainless steel is normally used.

4.4.4 Chemical Dosage Control

Control of chemical feeding is a critical part of any phosphorus removal system. Control of chemical dosages is important not only to ensure that effluent phosphorus requirements are consistently met, but also to keep chemical use and operating costs to a minimum. As discussed in Chapter 2, phosphorus loadings to a plant can be expected to fluctuate significantly on an hourly, daily, and even seasonal basis.

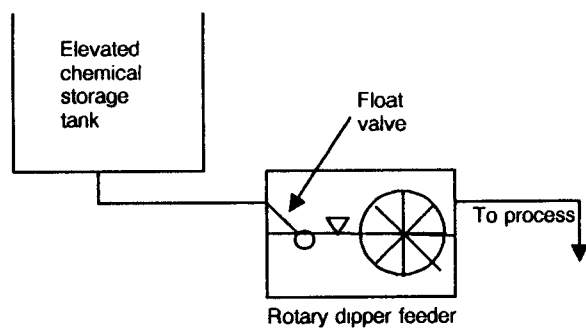
The type and complexity of selected control systems are dependent on plant size and sophistication of operations, as well as daily chemical usage. For small plants with low chemical requirements, a complex

automated control system cannot be justified on the basis of either economics or practicality. For large plants, however, close control of chemical dosage can result in substantial savings in chemical costs, and a more sophisticated control system would be within the operation and maintenance capabilities of the plant staff. Several control options are discussed below.

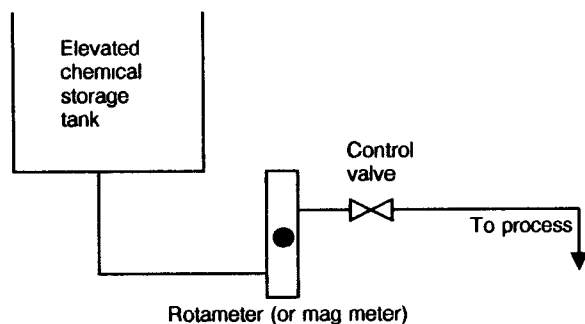
a. Manual Control

Manual dosage control may be appropriate for plants less than 3,785 m³/d (1 mgd). Manual operation of a chemical feed system would involve: 1) daily preparation of the chemical solution on a batch basis (if dry chemicals are purchased), and 2) manual setting of a control valve or the stroke of a diaphragm metering pump to establish proper flowrate of chemical solution. Although such a system is reliable and requires little maintenance, it has a major disadvantage in that chemical feed rate is constant

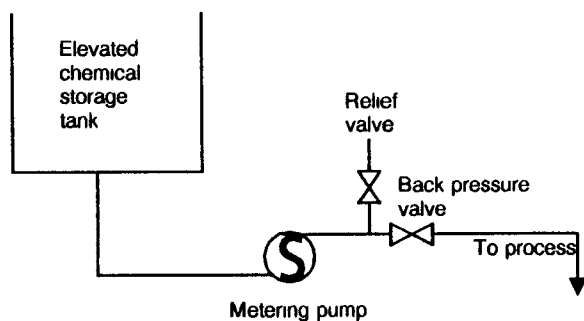
Figure 4-4. Liquid chemical feed alternatives for elevated storage.



a. Rotary dipper feeder



b. Rotameter with control valve



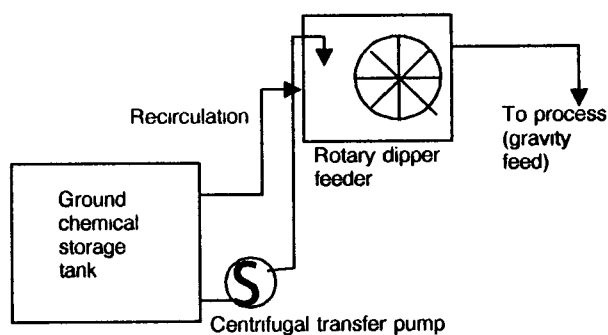
c. Metering pump

and plant flow is variable, resulting in variation in dosages. This problem can be overcome to some degree by manually varying the chemical feed rate at specified intervals based on historical flow and performance data (17). However, the potential for overdosing or underdosing exists. At plants where flows are generally predictable, this problem is minimal. At others, such as those with significant infiltration/inflow, operational schedules should reflect the need to adjust rates during initial periods.

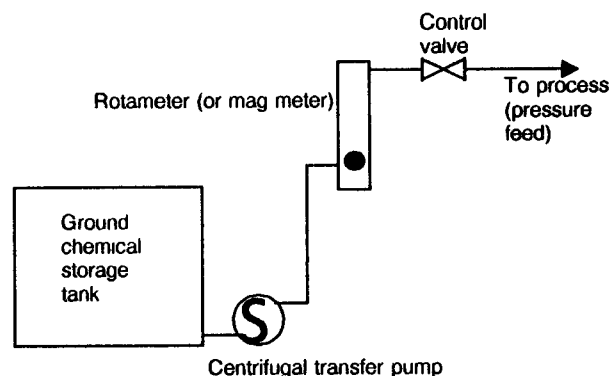
b. Flow Proportional and Programmed Control

Flow proportional control involves automatic adjustment of chemical feed rate in proportion to plant flow. This is accomplished by transmitting a signal

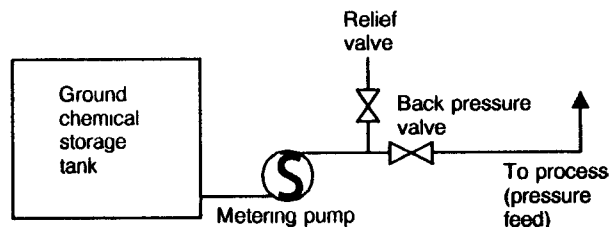
Figure 4-5. Liquid chemical feed alternatives for ground storage.



a. Rotary dipper feeder



b. Centrifugal pump with rotameter and control valve



c. Metering pump

from a flow measuring device to a controller, which then adjusts the speed of a rotary dipper feeder, the stroke of a metering pump, or the opening of a control valve. Such devices are well demonstrated and reliable for a variety of chemical feed applications. Unfortunately, the resulting chemical feed rate will be proportional to plant flowrate, which may bear little correlation to the actual phosphorus loadings to the facility.

An alternative to flow proportional control is programmed control in which chemical feed rate is programmed in relation to historical data on mass phosphorus loading patterns (product of plant flow and phosphorus concentration). The use of

programmable controllers is a relatively simple and effective technique which has been successfully implemented at full scale facilities. This requires an accurate knowledge of incoming phosphorus loadings, preferably on an hourly basis over a period of one week. Chemical feed rates are thus programmed to coincide with intervals of known phosphorus loadings. For automated systems, this approach is superior to flow proportional control since it provides for more efficient and specific application of chemicals. However, it still suffers from the fact that chemical dose is based on historical flow and phosphorus data. During wet weather flows, the programmed dosing schedule may not be applicable.

c. Feed Forward and Feedback Control

A more sophisticated approach for control of chemical dosage is the use of feed-forward or feedback control, in which the chemical feed rate is controlled in proportion to a combined signal of wastewater flowrate and some wastewater characteristic such as pH, conductivity, or phosphate concentration (16,18,19). In Norwegian studies, it was found that alkalinity was the predominant variable affecting alum dosage. Since field proven instruments that provided on-line analysis of alkalinity were not available, it was decided to base chemical dosage on conductivity measurements. Interestingly, good long-term correlation was found between alkalinity and conductivity. As a result, a conductivity feed forward, open loop control system was implemented to allow variation of chemical feed rate based on a combined signal of wastewater flowrate and primary effluent conductivity. This system was found to be effective, resulting in savings in chemical costs over other control alternatives. As of February, 1985, six plants in Norway using alum for phosphorus control had implemented the conductivity feed forward control system (18).

In-line phosphate analyzers have been investigated for use in automated chemical control systems (17). Unfortunately, operation and maintenance requirements are high, and skilled technicians must be available to service such units. Particularly when used for raw wastewater or primary effluent applications, the devices are subject to plugging and fouling. Overall, their use in chemical dosage control applications has been unsuccessful.

4.4.5 Chemical Mixing and Flocculation

It is important that chemicals used for phosphorus precipitation be intimately mixed with the wastewater to ensure uniform dispersion and efficient application of the chemicals. Where existing plants have been modified to remove phosphorus by chemical addition, it has been standard practice to use existing structures and facilities to provide mixing of the chemical with the wastewater. Points of addition have included Parshall flumes, drop manholes, aerated grit chambers, the discharge side of raw sewage pumps,

90° pipe bends, hydraulic jumps, and aeration basins, where the turbulence levels were typically higher than for other locations at the plant.

Although chemical addition at such locations has been reported to be effective, it is unlikely that optimum mixing conditions existed. Poor or inadequate mixing can result in inefficient chemical use and greater chemical consumption. Where viscous chemicals such as polymers are added to the waste stream, provision of intense mixing is essential. It has been recommended that polymer be added at a concentration of 0.01-0.05 percent, and never above 0.1 percent (14). This allows reduction in required mixing power by providing for improved dispersion upon injection into the main stream. Manufacturers' recommendations should be followed for polymer addition, since "overmixing" is possible, resulting in reduced performance.

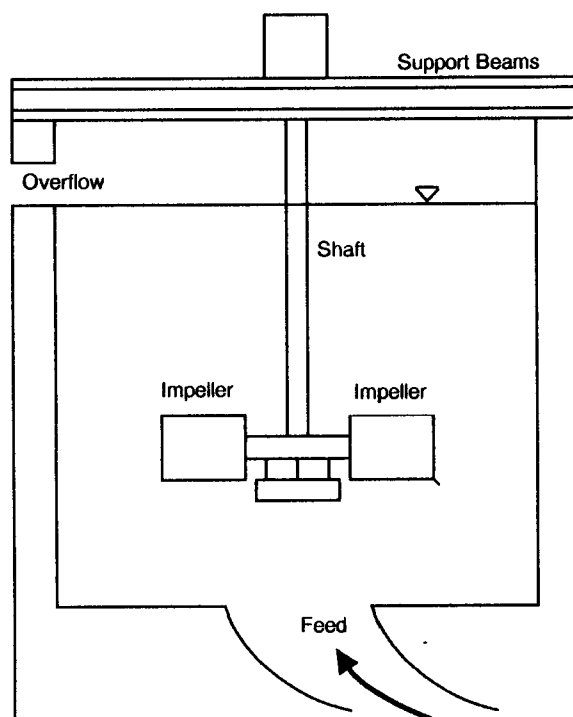
The most important parameters used in the design of mixing and flocculation processes are detention time, t , and velocity gradient, G . Velocity gradient is a measure of the shear intensity imparted to a fluid. The equation for velocity gradient is given in Section 4.5. For chemical mixing processes, detention times are approximately 30 sec, and G values are on the order of 300 m/sec/m although velocity gradients of up to 1,000 m/sec/m have been recommended (20). A typical flash mix tank is shown in Figure 4-6.

The flocculation process allows contact between coagulated solids so that they agglomerate to form solids with improved settling characteristics. Whether for a new plant or for a retrofit application, separate basins are seldom constructed specifically for flocculation. Rather, existing plant components such as aerated grit chambers, aerated distribution channels, or feed wells of clarifiers are used, often after some modification. Although use of existing tankage may be effective for flocculation, tank size and configuration should be carefully evaluated, and velocity gradients calculated to ensure presence of conditions that will promote good flocculation.

A common approach is the use of a flocculating clarifier, in which an expanded center well provides the desired detention time for flocculation. The contents of the flocculation well can be agitated by mechanical mixers or diffused air, although the hydraulic regime in the center well may be such that mechanical or air mixing does not provide additional benefit (21). A typical flocculating clarifier is shown in Figure 4-7.

Velocity gradients for flocculation processes generally are 20-80 m/sec/m, depending on the chemicals added and point of addition. Velocity gradients less than 50 m/sec/m may yield floc particles with too much trapped water, whereas velocity gradients above 80 m/sec/m may cause excessive floc shear

Figure 4-6. Typical flash mix tank.



and floc deterioration. In many cases, mineral salts are added directly to the aeration basin to allow mixing and flocculation to occur. Although this practice is effective, it represents less than ideal conditions for flocculation, as velocity gradients in aeration basins result in floc shear. In one study, air flowrates in the downstream end of the aeration basin were reduced to achieve a velocity gradient of 60 m/sec/m, which was found to be optimum for flocculation using ferric chloride (22). Addition of anionic polymer prior to clarification assists in the agglomeration of sheared floc.

Other parameters used in the design of flocculation processes are Gt , the product of velocity gradient and detention time, and GCt , where C is the ratio of volume of floc to total volume of suspension. In water treatment applications using aluminum or iron salts, Gt values are typically 30,000-150,000, with GCt values of 10-100 (23).

4.4.6 Clarification

Since clarifiers are used in virtually all wastewater treatment plants and are not specific to phosphorus removal systems, the following discussion will be limited to how their application or design differs from conventional practice when mineral salts are added to precipitate phosphorus.

The two most common points of mineral addition are: 1) prior to primary clarification and 2) prior to secondary clarification. Consequently, clarifier design

is critical to ensure adequate removal of suspended solids and consistent achievement of phosphorus discharge limitations.

Clarifiers used in chemical precipitation systems differ little from those employed in conventional biological treatment, although use of flocculation zones is recommended to allow flocculation to occur after addition of coagulants.

Provision of distinct flocculation zones is recommended for either primary or secondary clarifiers depending on the point of chemical addition. This is particularly important for primary clarifiers, since there may be little opportunity for flocculation to occur in existing processes.

For secondary mineral addition, flocculation can occur in aeration basins or channels preceding clarification, but the use of flocculation zones in secondary clarifiers is recommended practice to allow flexibility in the point of chemical addition and to provide a zone in which direct control can be exercised over velocity gradients in order to achieve optimum flocculation.

4.5 Design Methodology

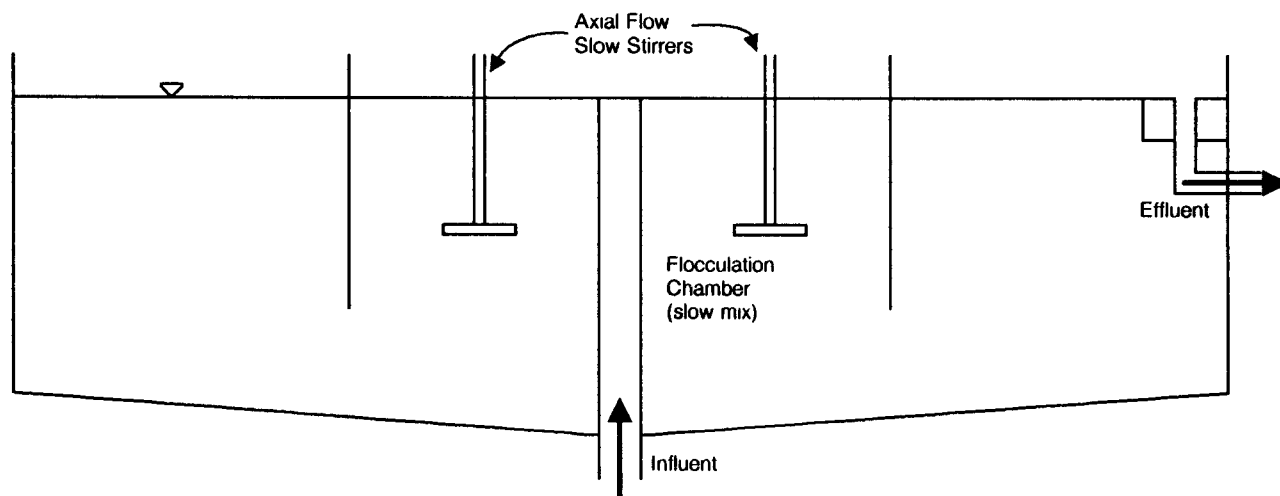
The approach to design of facilities for phosphorus removal using mineral salts follows the basic steps outlined below.

1. Characterization of the wastewater (raw wastewater, primary effluent, mixed liquor, secondary effluent).
2. Determination of chemical doses required to meet effluent phosphorus requirements based on addition to primary or secondary treatment.
3. Selection of chemicals and points of addition.
4. Evaluation of existing unit processes (retrofit applications).
5. Conduct of full-scale trials (retrofit applications).
6. Sizing and design of chemical storage, handling, and feed systems.
7. Sizing and design of liquid processes such as rapid mix and flocculation facilities, clarifiers, and aeration basins.
8. Sizing and design of sludge handling facilities.

4.5.1 Wastewater Characterization

A thorough knowledge of wastewater characteristics is required in order to proceed with chemical selection and process design. The following information is essential:

Figure 4-7. Typical mechanically mixed flocculating clarifier.



1. Average dry weather and wet weather flows.
2. Diurnal flow variation.
3. Characterization parameters.
 - a. BOD and COD
 - b. SS and VSS
 - c. Total and soluble phosphorus
 - d. Orthophosphate
 - e. pH
 - f. Alkalinity
4. Diurnal (hourly) variation in phosphorus loadings.

4.5.2 Determination of Chemical Doses

Jar tests should be conducted with the specific wastewater to be treated if possible. Such tests should be conducted over an extended period of time to obtain representative wastewater characteristics. Grab samples are collected at various times of the day and days of the week, and a range of dosages established for each chemical to be tested. A minimum of 10 data points should be gathered for each chemical dose on each waste stream investigated. Jar test procedures are described in Reference 24. Probability plots should then be developed which show percent of time that effluent phosphorus is less than a stated value for each dosage. These plots are valuable in determining the reliability of a particular coagulant over a range of dosages.

4.5.3 Selection of Chemicals

Chemical selection is made based on: 1) performance and reliability (results of jar tests) and 2) costs. Factors included in the selection decision include whether to purchase chemicals in dry or liquid form, in bulk or in small lots. Consideration should be given to whether secondary chemicals (e.g. polymers for

flocculation aids, bases for pH adjustment) will be required, and in what quantities.

When mineral salts are used for phosphorus precipitation, anionic polymers may be required, and prudent design should include polymer addition facilities. Adjustment of pH through addition of lime or other alkaline materials may be necessary, particularly with low-alkalinity wastewaters.

If waste products such as pickle liquor are available, jar testing should include investigation of such chemicals at various times to quantify variability of these chemicals. Also, the engineer should quantify their availability over long periods to determine storage requirements.

Assuming that more than one chemical or combination of chemicals yields acceptable results in terms of performance and reliability, the selection of chemicals is then an economic decision. A cost-effectiveness analysis should be conducted which addresses capital costs of chemical storage, handling, and feeding equipment, considering both dry chemicals and liquid chemicals in bulk and in small lots; as well as operational costs, including costs of chemicals, labor and power.

4.5.4 Evaluation of Existing Unit Processes

For retrofit applications the existing process train should be evaluated with respect to:

1. Potential points of chemical addition.
2. Degree of mixing at identified points of addition (e.g., raw sewage pumps, Parshall flumes or hydraulic jumps, aerated grit chambers, etc.).
3. Opportunities for flocculation (e.g., preaeration basins, clarifier feed wells, connection piping, weirs, aeration basins, etc.).

4. Adequacy of existing clarifiers (existing vs. required design overflow rates).

Potential points of chemical addition include: 1) prior to primary clarification, 2) prior to secondary clarification, 3) in aeration basin at turbulent locations, and 4) at several of these points simultaneously. When evaluating existing structures or equipment to provide mixing of the coagulant with the wastewater, velocity gradients should be calculated where possible to determine if the level of turbulence is sufficient for adequate mixing. A minimum velocity gradient of 300 m/sec/m is recommended. Canadian studies showed that efficient mixing could be obtained in full scale facilities by the following approaches (11):

1. Injection of the chemical into the discharge side of raw sewage pumps.
2. Chemical addition into a preaeration tank or aerated grit chamber.
3. Addition of chemical at a Parshall flume or similar turbulent constriction.
4. Chemical addition to pipe or channel between aeration tank and final clarifier with supplemental mechanical or air mixing at the point of addition.

If existing structures are inadequate for providing good mixing, separate mixing basins or modification of existing structures will be required. For example, existing preaeration basins can be modified to provide for intense mixing in the upstream end, followed by gentle mixing to promote flocculation. It is important to note that phosphorus removal efficiency can suffer significantly if intense mixing of the coagulant and wastewater is not provided. During the Canadian studies, it was found that, in several cases, phosphorus removal efficiency was doubled by increasing the intensity of mixing at the point of chemical addition (11).

In most cases, existing structures of the facility can be used, perhaps with some modification, to promote flocculation of the coagulated wastewater. Some flocculation will occur naturally during passage through the clarifier feed well, although consideration may be given to use of an enlarged center feed well in the clarifier to provide approximately 20 minutes of gently agitated flocculation.

The existing and design overflow rates in primary and secondary clarifiers must be compared with recommended designs for chemical precipitation systems. If necessary, additional clarification capacity should be provided. New primary clarifiers should be equipped with center flocculation zones. For secondary clarifiers preceded by chemical addition to the aeration basins, specially designed flocculation wells may be unnecessary, as air flowrates in the

downstream end of the aeration basins can be adjusted to provide near optimum conditions for flocculation.

4.5.5 Conduct of Full-Scale Trials

Where existing plants are to be retrofitted for phosphorus removal, full scale trials are recommended. Full-scale trials of 6 weeks for primary treatment plants and 8 weeks for secondary treatment plants have been used (25). Such trials offer the following advantages:

1. Determination of optimum point(s) of chemical addition.
2. Optimization of chemical dosages.
3. Evaluation of impacts on aeration requirements, SRT, F/M and other key process variables.
4. Direct observation of impacts of variables on effluent quality.
5. Estimation of the quantity (volumetric and mass) of additional sludge to be processed.
6. Observation and determination of impacts on sludge thickening, stabilization, and dewatering characteristics; chemical requirements for sludge conditioning; equipment modifications; and disposal limitations.

4.5.6 Design of Chemical Handling System

The main components of a chemical handling system include: 1) chemical storage facilities, 2) dry chemical feed and dissolution equipment (if dry chemicals are used), and 3) chemical solution feed systems.

a. Chemical Storage

Bulk storage facilities for either dry or liquid chemicals should be sized for capacities of at least 50 percent greater than the largest anticipated shipment, and should provide a minimum capacity of a ten day supply, and preferably a two week supply. Where chemicals are delivered by rail, it may be prudent to provide additional capacity to compensate for late or intermittent deliveries.

Dry chemical storage vessels should be designed with bottom hoppers having slopes of 60 degrees to help prevent bridging of the chemical. Vibrators are also used for this purpose. Dust collection equipment is recommended to control fugitive dust emissions during loading operations. Choice of materials is dependent on the particular chemical to be stored. For dry iron compounds, steel lined with rubber, asphalt, or plastic, synthetic resins, or concrete may be used. Dry alum can be stored in mild steel or concrete bins. Where dry chemicals are purchased in small lots such as bags or drums, the chemicals should be stored in enclosed, dry areas.

Liquid alum should be stored in tanks constructed of FRP, 316 stainless steel, or steel lined with rubber or PVC. Storage vessels for liquid sodium aluminate should be constructed of mild or stainless steel, FRP, or concrete. Use of copper and its alloys, rubber, and aluminum should be avoided. If located outside, tanks should be closed and vented, with heaters to maintain temperatures above -4°C (25°F). Polyurethane insulation between the tank and the pad is recommended.

Liquid iron solutions such as ferric chloride and ferrous chloride should be stored in vessels constructed of FRP, synthetic resins, or steel lined with rubber or plastic. Tank heaters and insulation may be required, as ferric chloride will crystallize at temperatures ranging from -12°C (10°F) for a 35 percent solution to 10°C (50°F) for a 45 percent solution. To prevent freezing during shipment, weaker solutions with lower freezing points are shipped during the winter.

Liquid chemical storage tanks should be equipped with a strip or float gauge to determine volume of tank contents, an access manhole, a filling hose with quick disconnect coupling, and a drain that is flush with the tank bottom.

b. Dry Chemical Feed and Dissolution Systems

Dry chemical feeder selection will be dependent on the chemical used. For dry alum, either open or enclosed construction is permissible. However, with dry iron compounds such as ferric or ferrous sulfate, belt type feeders are rarely used because of their open construction and exposure to corrosive vapors from the dissolving tank. Provision of water jets for vapor removal is recommended to protect equipment and operating personnel.

Selection of a volumetric vs. gravimetric feeder is dependent upon total chemical demand and degree of accuracy required. For plants larger than $3,785\text{ m}^3/\text{d}$ (1 mgd), gravimetric feeders are recommended, since the additional cost over a volumetric feeder is likely to be offset by the savings in chemical costs during the useful life of the equipment. Capacity ranges of feeders should be a minimum of 10 to 1. Sufficient capacity should exist to handle peak chemical demands. For some feeders, the capacity range can be significantly expanded by modification of the gear box or by use of variable speed drives.

For small plants using bagged chemicals, the feeder can be attached to a "day hopper," which is filled manually each day and which has a capacity for one day's supply of chemical. In larger plants, the chemical feeder is charged directly from the bulk storage silo.

Dry chemicals must be added to water in a dissolving tank to form a chemical solution prior to introduction

into the wastewater stream. For alum, dissolving tanks should be constructed of FRP, plastics, or 316 stainless steel. The most dilute alum solution recommended is a 6 percent solution (0.06 kg/l). For dry sodium aluminate, dissolving chambers may be mild steel or stainless steel. Preparation of a 6 percent solution is standard practice. Iron solutions are actively corrosive, so care must be taken in selecting materials. Ferric sulfate dissolvers should be constructed of type 316 stainless steel, plastics, rubber, or ceramics. Minimum solution strength for ferric sulfate is one percent, and the typical water to chemical weight ratio is 4:1. Ferrous sulfate dissolution tanks should be constructed of type 304 stainless steel, plastics, rubber, or iron. Recommended solution strength is 6 percent.

For most dry chemicals, dissolving chambers should be designed for a detention time of 5 minutes at the maximum feed rate. The exception to this is ferric sulfate, for which a 20-minute detention time is recommended (3). The dissolving tanks must be large enough to provide the necessary detention time at the maximum rate of feed of water and chemical. Dissolvers should be mechanically mixed and should be equipped with a water meter so that the proper solution strength can be determined and maintained.

c. Chemical Solution Feeding

As discussed in Section 4.4, rotary dipper feeders or metering pumps are commonly used to provide positive control over chemical feed rates. Other options are rotameters with control valves, or orifice plates. Flow control devices should be sized for the maximum expected flowrate of chemical solution.

Centrifugal pumps can be used for the transfer of chemical solutions from dissolving tanks to other vessels, or for direct feeding to the process through a flow measurement device such as a rotameter. To achieve uniform flow rates with centrifugal pumps, pumping head must be constant. Centrifugal pumps should be direct connected but not close-coupled to prevent leakage into the motor. Pump components for liquid alum service should be constructed of 316 stainless steel, FRP, or plastics. For ferric chloride, graphite or rubber lined pumps with Teflon seals are recommended.

Metering pumps are typically of the positive displacement type, either diaphragm or plunger. Diaphragm pumps protected with internal or external relief valves are preferred. A back pressure valve is recommended to provide positive check valve operation (3).

Materials of construction for chemical feed service include 316 stainless steel, FRP, plastics, and rubber. Manufacturers' recommendations should be followed regarding selection of pump materials for the specific chemical of choice.

For pipes transporting alum solution, use of FRP, PVC, or other plastics is recommended. Valves should be plastic, 316 stainless steel, or rubber-lined iron or steel.

For ferric chloride conveyance, pipes should be constructed of steel lined with rubber or Saran, FRP, or plastics. Valves should be rubber- or resin-lined diaphragm valves, Saran-lined valves with Teflon diaphragms, rubber-sleeved pinch valves, or plastic ball valves (3,4).

Pipe selection for polymer service should be made after the type of polymer has been determined. Plastic pipe or type 316 stainless steel is normally used.

4.5.7 Design of Liquid Processes

Design of chemical phosphorus precipitation systems requires consideration of not only chemical storage and feed equipment, but also mainstream process equipment such as aeration basins and clarifiers as well as sludge handling equipment used for thickening, stabilization, dewatering and disposal. For example, where phosphorus is removed by chemical precipitation in primary treatment, primary clarifier design will likely vary from standard practice to provide for flocculation. Since BOD and SS removal efficiencies will be significantly improved, loadings to downstream biological processes will be reduced, while sludge generation rate will be increased. In addition, sludge characteristics may be considerably different than if no chemicals were added.

a. Rapid Mixing

In the design of new facilities incorporating chemical phosphorus precipitation, provision of separate rapid mixing basin(s) is recommended for complete contact of the chemical with the wastewater. Rapid mix tanks may be designed for detention times of 20-60 seconds, although 30 seconds is recommended for metal salts. Velocity gradients (G) on the order of 300 m/sec/m are generally sufficient, although values as high as 1,000 m/sec/m have been recommended (15). The power required to maintain turbulent conditions ($Reynolds > 10^5$) in a flash mix basin with an impeller mixer can be calculated from (26):

$$P = \rho K_T n^3 D_a^5 / g$$

where,

- P = power requirement, ft-lb/sec
- ρ = mass density of the fluid, lb/ft³
- n = impeller revolutions per second, rps
- D_a = diameter of impeller, ft
- g = acceleration due to gravity, 32.2 ft/sec²
- K_T = constant

Values of K_T for various mixing devices are shown in Table 4-7).

Table 4-7. Values of K_T for Determining Impeller Power Requirements (26)

Type of Impeller	K_T
Propeller (square pitch, 3 blades)	0.32
Propeller (pitch of 2, 3 blades)	1.00
Turbine (6 flat blades)	6.30
Turbine (6 curved blades)	4.80
Turbine (6 arrowhead blades)	4.00
Fan turbine (6 blades)	1.65
Flat paddle (2 blades)	1.70
Shrouded turbine (6 curved blades)	1.08
Shrouded turbine (with stator, no baffles)	1.12

Velocity gradient can be calculated from the general equation (26):

$$G = [P/V\mu]^{1/2}$$

where,

- G = velocity gradient, ft/sec/ft
- μ = absolute fluid viscosity, lb-sec/ft²;
approximately 2×10^{-5} lb-sec/ft² @ 20°C
- V = basin volume, ft³

For an electrically driven mechanical mixer,

$$P = (WHP)(550)$$

where,

- WHP = delivered water horsepower or
(KVA)x(Motor Eff.)x(Power Factor)/0.746
- KVA = apparent power, kVA

b. Flocculation

As discussed in Section 4.4.5, numerous devices have been used to promote flocculation, including preaeration basins, interprocess channels, aeration basins, and enlarged feed wells of clarifiers. Maintenance of velocity gradients in the range of 50 to 80 m/sec/m for no more than 15 minutes is recommended for good flocculation where metal salts are used. Some have recommended high energy flocculation for 5 minutes followed by gentle flocculation for 15 minutes (15). Such "tapered flocculation" can occur naturally in clarifiers and in aeration basins in which air flows are decreased at the downstream end.

The degree of flocculation is difficult to predict, since it will vary with wastewater characteristics and choice of chemicals, as well as with hydraulic characteristics and energy dissipation in the basin. For this reason, design of new facilities should provide for maximum flexibility in order to achieve optimum conditions for flocculation.

Flocculation can be accomplished by mechanical (paddle or turbine) mixers, air diffusers, or baffles. Velocity gradient can be determined by substituting the appropriate value of P into the general equation below:

$$G = [P/V\mu]^{1/2}$$

Paddles:

$$P = C_D \omega A v^3 / 2g$$

where,

- P = power requirement, ft-lb/sec
- C_D = coefficient of drag of flocculator paddles, dimensionless
- ω = mass fluid density, slugs/ft³
- A = area of paddles, ft²
- v = relative velocity of paddles in fluid, ft/sec (about 0.7-0.8 of paddle tip speed)

Turbines:

$$P = K_T \omega D_a^5 v^3 / g$$

where,

- P = power requirement, ft-lb/sec
- K_T = constant
- D_a = diameter of impeller, ft

Air:

$$P = 82 Q_a \log [(H + 34)/34]$$

where,

- P = power imparted to water, ft-lb/sec
- Q_a = air supplied, scfm
- H = head of water above air diffusers, ft

Baffles:

$$P = Q \omega h_f$$

where,

- P = power imparted to water, ft-lb/sec
- Q = flow, ft³/sec
- h_f = head loss due to friction, ft

Important considerations in the design of flocculation basins include:

1. Transport conditions -- minimize transport time (<1 minute) from rapid mix tank to flocculator. Keep velocities low and avoid turbulence from flocculator to clarifier.
2. Flow distribution -- maintain good flow distribution at flocculator inlet by designing for

headloss through inlet ports of 10 times transport headloss from first to last port.

3. Short circuiting -- install baffle walls between multiple paddle units in rectangular basins; in circular units with rotary flocculators, place baffles along wall to prevent entire tank contents from rotating with mechanism.
4. Tapered flocculation -- provide compartments which increase in volume from the inlet end to the outlet end of the tank. Reduce slow mix power with each downstream stage.
5. Separate vs. combined units -- consider combined flocculator-clarifier for lime precipitation systems; compare costs, reliability, flexibility of separate vs. combined unit processes.

When flocculation is carried out in an aeration basin or separate structure, the velocity in conduits conveying the floc to the clarifier should be kept between 0.15 and 0.30 m/sec (0.5-1.0 ft/sec) so as to prevent destruction of the floc (8).

c. Clarification

If not preceded by a flocculation process, clarifiers receiving a chemically dosed effluent should be designed with a flocculation zone. Several manufacturers provide circular clarifiers with center flocculation wells which incorporate a detention time of 20-30 minutes. Generally, flocculation wells are 30-40 percent of the tank diameter. The walls of the feedwell generally extend down to 60-75 percent of the tank depth (14).

The principal design criterion for chemical clarification is overflow rate. Table 4-8 provides a summary of recommended design overflow rates for primary, secondary, and tertiary clarifiers receiving wastewater coagulated with mineral salts (14). These values are generally conservative.

Minimum bottom slopes of clarifiers should be 8 percent. All clarifiers should be equipped with scum removal mechanisms. A minimum sidewater depth of 3.6 m (12 ft) is recommended (14).

4.5.8 Design of Sludge Handling System

Addition of mineral salts for phosphorus precipitation may significantly increase the quantity of sludge generated due to the production of metal-phosphate precipitates and metal hydroxides as well as the improved removal of suspended solids.

Addition of metals upstream of the primary clarifier will result in a primary sludge mass increase of 50-100 percent. This increase is normally due almost equally to improved capture of suspended solids and additional chemical sludge (3). Overall plant sludge

Table 4-8. Recommended Overflow Rates for Conventional Clarifiers Receiving Wastewater Coagulated with Mineral Salts (14)

Type of Clarifier	Design Overflow Rates (Average Flow)	
	w/o Polymer m ³ /m ² /d (gpd/ft ²)	w/Polymer m ³ /m ² /d (gpd/ft ²)
Primary	24 (600)	49 (1,200)
Secondary	24 (600)	-
Tertiary	24 (600)	49 (1,200)

mass increase is much smaller owing to reduced secondary sludge production from improved primary removals, e.g., 60-70 percent increase across the entire plant.

For metal addition to secondary processes, waste mixed liquor sludge mass can be expected to increase by 35-45 percent. Overall plant sludge mass increase will be 5-25 percent (11). Metal addition to either primary or secondary treatment will not only increase sludge mass, but sludge volume will be increased since settled sludge concentration in the clarifiers may decrease by up to 20 percent (11). A detailed discussion of characteristics, generation rates, and treatment alternatives for sludges resulting from chemical phosphorus removal systems is presented in Chapter 5.

4.5.9 Design Example

Develop a preliminary process design for retrofitting an existing 11,350 m³/d (3.0 mgd) activated sludge facility for chemical precipitation of phosphorus. The following effluent requirements are imposed:

BOD = 15 mg/l
SS = 15 mg/l
Total P = 0.8 mg/l

1. Characterize wastewater

BOD = 200 mg/l
SS = 210 mg/l
Total P = 7 mg/l
Ortho P = 4 mg/l
Alkalinity = 220 mg/l
pH = 7.1
Design ave. daily dry weather flow = 11,350 m³/d
Current ave. daily dry weather flow = 9,500 m³/d
Peak:average daily flow ratio = 2.5:1

2. Determine chemical dosages

Conduct jar tests using samples of both raw wastewater and mixed liquor. Raw wastewater

samples should be collected at various times of the day and on various days of the week. Mixed liquor samples should be collected at different operating conditions (e.g. SRT) where possible. Evaluate effectiveness of coagulants including ferric chloride, alum, and ferrous sulfate over a prescribed range of dosages. If waste pickle liquor is available locally or at a reasonable cost, include in jar tests.

Plot effluent total phosphorus vs. chemical dose on a probability plot for each chemical. Determine dosages that will meet effluent phosphorus requirements at the desired level of probability (e.g., 90 percent or 95 percent).

3. Select chemical

Chemicals should be selected based on the following criteria:

- Performance and reliability (from probability plots)
- Unit costs
- Reliability of supply
- Operation and maintenance of chemical feed equipment
- Safety in handling

Based on analysis of the above criteria, assume ferric chloride is selected as the most economical and reliable chemical.

4. Evaluate existing liquid stream processes

An evaluation of existing liquid stream processes resulted in the following observations and conclusions:

a. Potential points of chemical addition

Two identified points of possible chemical addition to the existing activated sludge facility are: 1) upstream of the primary clarifiers and 2) in or immediately after the aeration basin.

b. Degree of mixing

Inspection and evaluation of hydraulic structures upstream of primary clarification indicate that sufficient mixing and turbulence do not exist, thus requiring construction of a separate rapid mix basin. For chemical addition to secondary treatment, it is concluded that direct chemical addition to the aeration basins is feasible without modification of the basins.

c. Opportunities for flocculation

Evaluation of the hydraulic characteristics of the existing primary clarifiers indicates that adequate flocculation will occur in the feed well and other portions of the clarifier. This will be confirmed in full-scale trials.

For addition to the aeration basin, it is determined that air flowrates through the diffusers in the downstream portions of the basins can be readily throttled to yield desired velocity gradients for good flocculation. Existing velocity gradients are approximately 120 m/sec/m. Full-scale trials are necessary to determine optimum velocity gradients in the aeration basin.

d. Adequacy of existing clarifiers

The existing facility has 2 primary clarifiers, each 15 m (50 ft) in diameter. Design overflow rate at 11,350 m³/d (3 mgd) average daily design flow is 32 m³/d/m² (800 gpd/ft²), and approximately 27 m³/d/m² (670 gpd/ft²) at present raw wastewater flow. Comparison with design overflow rates for chemical precipitation systems indicates that the current overflow rate is only about 10 percent greater than recommended [24 m³/d/m² (600 gpd/ft²)]. Since the recommended rates are conservative, expansion of existing clarifier hydraulic capacity is not warranted at this time.

The existing final clarifiers were designed for an overflow rate of 24 m³/d/m² (600 gpd/ft²) at average flow. Current overflow rates are approximately 20 m³/d/m² (500 gpd/ft²). These rates are within recommended guidelines of 24 m³/d/m² (600 gpd/ft²) for chemical clarification. Thus, expansion of secondary clarifier hydraulic capacity is not justified.

5. Conduct full-scale trials

Full-scale trials were conducted over period of eight weeks using ferric chloride as a coagulant. Key findings from the studies are as follows:

- a. Split addition of ferric chloride to primary and secondary treatment produced optimum results.
- b. Optimum dosages were 8 mg/l to primary treatment and 32 mg/l to secondary.
- c. Addition of a rapid mix basin ahead of primary treatment is necessary. For the trials, temporary mixing facilities were constructed.
- d. Good flocculation occurs naturally prior to the primary clarifier.
- e. Injection of ferric chloride at a point halfway down the aeration basin was found to be

effective when air flows in the downstream third of the basin were reduced to yield velocity gradients of approximately 70 m/sec/m.

- f. Effluent suspended solids were generally less than 10 mg/l.
- g. Effluent total phosphorus concentrations ranged from 0.5 to 0.8 mg/l.
- h. Modification of existing clarifiers is unnecessary.

6. Design chemical handling system

From full-scale trials, it was found that split addition of ferric chloride to primary and secondary treatment was the most economical and reliable approach, with 20 percent of the chemical added upstream of the primary clarifiers and 80 percent added directly to the aeration basins. Optimum dosages were approximately 8 mg/l and 32 mg/l to primary and secondary treatment, respectively. Ferric chloride is available locally in liquid form.

Total FeCl₃ requirements at average design flow:

$$40 \text{ mg/l} \times 11,350 \text{ m}^3/\text{d} \times 0.001 = 454 \text{ kg/d}$$

Average volumetric requirements for 40% FeCl₃ solution:

$$(454 \text{ kg/d}) \div (0.40 \times 1,400 \text{ kg/m}^3) = 0.81 \text{ m}^3/\text{d} \quad (215 \text{ gal/d})$$

Peak volumetric requirements:

$$2.5 \times 0.81 \text{ m}^3/\text{d} = 2.02 \text{ m}^3/\text{d} \quad (535 \text{ gal/d})$$

Size ferric chloride storage tanks based on 1-1/2 times largest anticipated shipment or 10 days storage at maximum feed rate:

$$1.5 \times 15 \text{ m}^3 = 22.5 \text{ m}^3 \quad (5,940 \text{ gal})$$

$$10 \text{ days} \times 2.02 \text{ m}^3/\text{d} = 20.2 \text{ m}^3 \quad (5,340 \text{ gal})$$

Use 2 storage tanks for total of 23 m³ (6,000 gal) storage capacity.

Select 2 metering pumps (1 standby); positive displacement diaphragm type. Size each pump at twice the anticipated maximum chemical feed rate - 85 l/hr (22.3 gph). Since data collected indicate a relatively constant diurnal phosphorus concentration in the raw wastewater, design for flow proportional control of chemical metering pumps.

7. Design liquid stream processes

Full-scale trials indicate that no existing hydraulic components upstream of primary treatment provide

sufficient turbulence to achieve adequate mixing. Therefore, a separate rapid mix tank is required upstream of the primary clarifiers. Existing primary and secondary clarifiers do not require modification to achieve desired performance.

Rapid Mix Tank

Size for detention time of 30 sec at average flow:

Volume = 3.3 m³ (870 gal)

Dimensions = 1.5 m x 1.5 m x 1.5 m deep

Use propeller mixer with $K_T = 1.00$ (Table 4-7)

Power required for mixing:

$$P = \rho K_T n^3 D_a^5 / g$$

Use $\rho = 62.4 \text{ lb/ft}^3$

$K_T = 1.00$

$n = 600 \text{ rpm (10 rps)}$

$D_a = 1 \text{ ft}$

$g = 32.2 \text{ ft/sec}^2$

$$\begin{aligned} P &= 62.4(1.00)(10)^3(1)^5/32.2 \\ &= 1,940 \text{ ft-lb/sec} \\ &= 1,940 \div 550 = 3.5 \text{ hp (use 4 hp)} \end{aligned}$$

Check velocity gradient, G :

$$\begin{aligned} G &= [P/V\mu]^{1/2} \\ &= [(4 \times 550) \div (3.3 \times 35.31)(2 \times 10^{-5})]^{1/2} \\ &= 974 \text{ sec}^{-1} \text{ (within the range discussed in} \\ &\quad \text{Section 4.5.7)} \end{aligned}$$

8. Design sludge stream processes

Although techniques are available to calculate additional sludge production from chemical addition (see Chapter 5), the most reliable method is through full-scale trials. In addition, such trials allow determination of the impact of chemical addition on sludge thickening, stabilization, and dewatering characteristics. A design example for sludge handling in chemical phosphorus removal systems is presented in Chapter 5.

4.6 Retrofit Considerations

In general, existing wastewater treatment plants can be retrofitted for chemical phosphorus precipitation relatively easily and inexpensively, provided that sufficient hydraulic and solids handling capacity is available. Design considerations include:

1. Existing vs. design flows and loadings (including phosphorus); average and peak.
2. Existing and design overflow rates for primary and secondary clarifiers, compared to recommended values for chemical precipitation systems.

3. Current loadings to secondary biological processes.

4. Solids loadings to sludge thickeners and dewatering equipment, compared to expected loadings with chemical precipitation.

5. Impacts of chemical addition on sludge thickening, stabilization, and dewatering characteristics.

6. Impacts on volume of sludge for final disposal.

7. Availability of turbulent zones for chemical addition points.

8. Availability of gently mixed zones for flocculation of chemically coagulated wastewater.

The two variables which have the greatest impact on the cost and complexity of retrofitting plants for chemical phosphorus removal are the existing solids handling and clarification capacity.

For any chemical precipitation system, sludge production can be expected to increase substantially, requiring a thorough capacity analysis of sludge handling processes and equipment. In some cases, existing equipment may be sufficient, requiring only longer operating times for such sludge processing operations such as dewatering. In other situations, expansion of existing processing capability through construction of additional tankage or installation of new equipment may be necessary, involving potentially large capital investments.

Design of clarifiers for precipitation of phosphorus by mineral addition is typically conservative, although when polymer is used as a flocculant aid, allowable hydraulic loadings can often be significantly increased without degradation of effluent quality.

In order to achieve low levels of phosphorus in plant effluents, suspended solids concentrations must be correspondingly low. It is generally true that, unless effluent suspended solids concentrations can be reduced to below about 15 mg/l, it is impossible to achieve an effluent total phosphorus concentration of less than 1 mg/l, even though soluble phosphorus may be as low as 0.1 mg/l (11).

4.7 Case Histories

4.7.1 Orillia, Ontario

This case history is included to document the capabilities of a dual point chemical addition system to achieve low effluent phosphorus levels. The information has been extracted from the final report of a study funded by the Ontario Ministry of the Environment (27).

The Orillia, Ontario wastewater treatment plant employs the conventional activated sludge process. Unit operations include preliminary treatment, primary clarification, aeration, secondary clarification, and chlorination. Primary and secondary sludge is pumped to a two-stage anaerobic digestion system. The digested sludge is discharged to lagoons for storage during winter months, and spread on agricultural land during other times of the year. Phosphorus removal is normally practiced by adding alum to the outlet of the aeration tanks at a dosage of approximately 65 mg/l. Target effluent phosphorus levels are 1 mg/l. Design flow is 18,000 m³/d (4.8 mgd). Plant performance data are summarized in Table 4-9.

Table 4-9. Summary of Plant Data During Single-Point and Dual-Point Alum Addition (27)

Parameter	Single-Point Alum Addition	Dual-Point Alum Addition
Flow, m ³ /d	14,000	17,000
Raw sewage		
BOD, mg/l	115	138
SS, mg/l	197	182
T-P, mg/l	8.4	5.2
Soluble P, mg/l	2.9	1.8
Primary effluent		
BOD, mg/l	62	52
SS, mg/l	89	64
T-P, mg/l	4.5	2.7
Soluble P, mg/l	2.0	0.9
Final effluent, mg/l		
BOD, mg/l	11	17
SS, mg/l	10	14
T-P, mg/l	0.65	0.36
Soluble P, mg/l	0.18	0.06
Alum dosage, mg/l		
Primary process	0	16
Secondary	64	32
mg/l Al ³⁺ + /mg/l P rem.	1.3	1.1

A study was initiated in 1981 to assess the feasibility of achieving effluent total phosphorus concentrations of 0.3 mg/l. It was originally intended that this would be achieved through chemical addition plus tertiary filtration. However, initial jar tests indicated that the effluent P objective of 0.3 mg/l might be obtained by dual point chemical addition without tertiary filters. The first phase of the study involved optimizing existing operations. To improve mixing and flocculation, the point of chemical addition was moved from a manhole between the aeration basins and final clarifiers to a point near the last air diffuser in the aeration basin. This allowed reduction in the alum dosage to 60 mg/l.

Full-scale performance of dual point alum addition for phosphorus removal was then investigated. The

scheme involved simultaneous addition of approximately 20 mg/l of alum to the raw wastewater at the aerated grit chamber and approximately 40 mg/l of alum to the tail end of the aeration basin. A sampling program was conducted during the period September 1981 through June 1982 to evaluate the impact of dual chemical dosage points. Data collected during this period are also summarized in Table 4-9.

Comparison of data for the two periods shows that the use of dual point chemical addition improved the removal of phosphorus significantly without increasing the overall alum dosage. Average effluent P of 0.36 mg/l was achieved at an overall dosage of 48 mg/l during dual point addition, whereas a dosage of 64 mg/l to the secondary process alone only achieved an effluent P of 0.65 mg/l. With dual point addition, average soluble P in the effluent was 0.06 mg/l, vs. 0.18 mg/l with single point chemical addition. Most of the phosphorus leaving the system during either period was associated with suspended solids. Note that effluent SS and BOD were somewhat higher during this period, possibly due to the fact that the average wastewater flow was 21 percent greater. Secondary clarifier average overflow rates were also higher at 35 m³/d/m² (860 gpd/ft²) vs. 30 m³/d/m² (750 gpd/ft²) during single point addition. It was concluded that the 0.3 mg/l P standard could be consistently met with dual point, flow-paced chemical addition with no tertiary filtration. Clarifier overflow rates are critical in maintaining low effluent SS and total P levels.

4.7.2 Elizabethtown, Pennsylvania (13)

The wastewater treatment plant serving the City of Elizabethtown consists of screening, grit removal, flow equalization, primary clarification, two stage trickling filter system, rapid mix chambers, polymer mix chambers, final clarification using clariflocculators, and chlorination. Primary and secondary sludge is blended, thickened, anaerobically digested, and applied to agricultural land. Alum or ferric chloride is added to the rapid mix chamber for phosphorus removal. Polymer can be added immediately downstream of the rapid mix chamber in a separate mixing chamber. The NPDES permit requires the plant to achieve the following effluent quality: BOD - 30 mg/l, SS - 30 mg/l, and total P - 2 mg/l.

Table 4-10 summarizes the performance of the plant from July, 1985 through April, 1986. Alum and polymer were used for phosphorus removal through January of 1986, at which time plant personnel began using ferric chloride because of its anticipated lower cost. When ferric chloride addition began, polymer was no longer used.

During the period of alum addition, average alum dosage was 9.9 mg/l as Al, and average polymer dosage was 0.4 mg/l. The ratio of aluminum applied to phosphorus removed (Al:P) was 1.35. When alum

Table 4-10. Summary of Performance at Elizabethtown, PA (13)

Month	30-Day Average Values						
	BOD			SS		P	
	Flow	In	Out	In	Out	In	Out
	m ³ /d	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
July, 1985 ¹	5,870	157	7	189	14	7.0 ³	2.2
August ¹	5,830	175	18	220	32	7.0 ³	2.0
September ¹	6,130	140	8	202	10	7.0 ³	0.7
October ¹	6,020	142	6	243	9	9.5	0.9
November ¹	6,960	106	3	279	12	8.1	0.9
December ¹	7,420	91	4	249	15	6.0	0.8
January, 1986 ¹	7,270	120	5	179	14	6.6	1.0
February ²	10,600	66	7	119	16	3.6	1.4
March ²	8,590	88	6	171	15	6.0	1.9
April ²	7,000	110	8	207	12	5.6	1.8
Ave. July-Jan ¹	6,510	133	7	223	15	7.3	1.2
Ave. Feb-April ²	8,740	88	7	166	14	5.1	1.7

¹ Alum and polymer used for P removal. Alum dose = 9.9 mg/l as Al (Al:P = 1.35). Polymer dose = 0.4 mg/l.

² Ferric chloride used for P removal. FeCl₃ dose = 7.1 mg/l as Fe (Fe:P = 1.39).

³ Estimated values.

and polymer were replaced by ferric chloride in February of 1986, average ferric chloride dosage was 7.1 mg/l as Fe, for an Al:P ratio of 1.39. Plant performance was similar with alum and polymer vs. ferric chloride. Concentrations of BOD, SS, and P in the plant effluent were well within discharge requirements.

Data on sludge production (dry solids basis) were not available. It was estimated that approximately 40 percent of the sludge was chemical sludge when alum and polymer were used, and approximately 30 percent of the sludge was chemical sludge when ferric chloride was used.

Estimated unit costs for phosphorus removal, including chemicals and chemical sludge disposal were \$3.52/kg (\$1.60/lb) of P removed with alum and polymer, and \$2.73/kg (\$1.24/lb) of P removed using ferric chloride (1986 costs).

4.7.3 Little Hunting Creek, Virginia (13)

The Little Hunting Creek wastewater treatment plant provides advanced secondary treatment using single-stage trickling filters. Unit processes consist of raw wastewater screens, grit chamber, primary clarifiers, trickling filters, secondary clarifiers, and chlorine contact tanks. Secondary sludge is returned to the headworks for co-settling with primary sludge. Waste sludge is conditioned with lime, dewatered by vacuum filters, and hauled to the Lower Potomac Pollution Control Plant for incineration. Phosphorus is removed by ferric chloride addition to a manhole at the headworks, and polymer addition: 1) just

downstream of the Parshall flume and 2) between the trickling filters and final clarifiers. Effluent discharge goals are 20 mg/l BOD, 20 mg/l SS, and 0.2 mg/l P (voluntary).

Performance data for the period July 1984 through June 1985 are shown in Table 4-11. The plant effluent was consistently within NPDES permit requirements for BOD and SS. Based on 30-day average values, the plant met the voluntary effluent P limit of 0.2 mg/l for nine of the twelve months.

Chemical requirements for the plant are high. Ferric chloride dosage is approximately 125 mg/l (43 mg/l as Fe), corresponding to a Fe:P ratio of 4.7. Total dosage of polymer (two points of addition) is approximately 2.8 mg/l. The high dosages are indicative of difficulties involved in retrofitting the plant for phosphorus removal.

Physical restrictions which prevent reduction of chemical dosages include: 1) limited points for chemical addition, 2) non-ideal conditions for mixing and flocculation, 3) absence of a tertiary filter, and 4) inability to accurately split the flow between the two trickling filters. Dual point chemical addition is not possible because the shallow depths of the secondary clarifiers are insufficient to handle excess chemical sludge.

Data on sludge production were not collected. It was estimated that approximately 44 percent of the sludge handled was chemical sludge from phosphorus removal.

Table 4-11. Summary of Performance at Little Hunting Creek, VA (13)

Month	30-Day Average Values						
	BOD			SS		P	
	Flow	In	Out	In	Out	In	Out
	m ³ /d	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
July, 1984	16,200	169	10	190	8.6	9.9	0.22
August	15,200	167	10	183	6.6	10.0	0.16
September	14,100	183	8	200	7.6	9.4	0.15
October	13,900	180	8	188	7.0	10.1	0.16
November	13,700	186	11	191	6.5	9.7	0.15
December	13,900	196	11	169	6.8	10.1	0.17
January, 1985	14,100	202	11	197	8.1	9.0	0.20
February	17,300	153	14	147	9.3	8.3	0.22
March	14,500	160	14	149	8.8	8.9	0.19
April	13,400	164	11	149	9.3	8.6	0.16
May	13,600	162	9	179	10.0	8.4	0.20
June	13,200	169	13	177	10.3	9.2	0.26
Average	14,400	175	11	177	8.3	9.3	0.19

The unit cost of phosphorus removal including chemicals, sludge conditioning, and chemical sludge handling was estimated to be \$13.39/kg P removed (\$6.08/lb P). Approximately 68 percent of this cost was attributed to sludge handling.

Table 4-12. Cost Estimates for Phosphorus Removal by Mineral Addition.

Flow Rate	Chemical	Chemical Dose ^a	Installed Equipment Cost ^{b,c}	Annual O&M Cost ^{b,d}
m ³ /d		mg/l	\$	\$/yr
3,785	Liquid alum	65	75,600 ^d	44,000
	Liquid FeCl ₃	50	65,700	40,000
37,850	Liquid alum	65	122,800	328,000
	Liquid FeCl ₃	50	116,800	369,000

^a Alum as Al₂(SO₄)₃•14H₂O

Dosages are average values from Reference 23 for mineral addition to secondary processes to meet effluent TP = 1.0 mg/l.

^b Cost estimates from EPA report, *Costs of Chemical Clarification of Wastewater*, 1976, updated to May 1987.

^c Includes chemical storage and transfer facilities, chemical feed pumps and piping, and rapid mix basin. Does not include additional sludge handling equipment.

^d Includes O&M costs for labor, power @ \$0.06/kWh, maintenance materials and chemicals. Does not include O&M for handling additional sludge generated. Chemical costs based on:

Liquid alum in bulk: \$0.17/kg

Liquid FeCl₃ in bulk: \$0.20/kg

4.8 Costs

Table 4-12 summarizes cost estimates for a chemical addition system for phosphorus removal. Capital and operation and maintenance costs are presented for chemical feed systems for 3,785-m³/d (1-mgd) and 37,850-m³/d (10-mgd) plants. Sludge handling costs are not included. These costs are planning level cost estimates (± 30 percent).

4.9 References

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

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

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CHAPTER 5

Sludge Handling

5.1 Introduction

The importance of sludge handling in the design of new facilities or in the retrofitting of existing plants for phosphorus removal cannot be over emphasized. The difficulty in predicting generation rates and characteristics of sludges derived from phosphorus removal operations is further complicated by the wide variability of data and the inconsistent information reported from operating phosphorus removal facilities.

Retrofitting of existing facilities to achieve phosphorus removal presents difficult design challenges with respect to sludge handling. For example, addition of chemicals to precipitate phosphorus from the liquid stream is relatively straightforward and can be implemented with relatively small capital expenditure. However, the additional sludge generated and the impact on sludge characteristics can easily result in overloading of existing thickening, stabilization, and dewatering equipment, or require significant increases in operational staff time to operate and maintain this equipment. Installation of new capital-intensive equipment may be required to effectively process the additional sludge. On the other hand, at some facilities, implementation of a phosphorus removal system may not significantly impact the existing sludge handling operation. Such variability of experience with sludge handling at plants removing phosphorus makes development of firm design recommendations difficult.

The objectives of this chapter are to:

1. Present information on the quantities and characteristics of sludges derived from chemical and biological phosphorus removal processes.
2. Provide specific recommendations for the design of sludge handling processes and equipment used in plants removing phosphorus.
3. Summarize the experience with sludge handling processes at full-scale phosphorus removal facilities.

It should be noted that, in some cases, specific design criteria cannot be presented because of the

wide variability in sludge characteristics, resulting in variations in performance of sludge handling processes and equipment. In these instances, full-scale experience is summarized and a design approach is outlined.

5.2 Current Practice for Handling Chemical Sludges

5.2.1 Introduction

In 1976 and 1977, an extensive EPA-sponsored survey was conducted of 174 municipal wastewater treatment plants in the United States and Canada which used chemicals to remove phosphorus (1). The purpose of this survey was to quantify the effects of chemical addition on the sludge handling and disposal operations at full-scale plants.

This section summarizes the results of this survey with regard to the types of facilities, points of chemical addition, sludge generation rates, and prevalence of various sludge treatment and disposal methods. Information regarding the impacts of chemical addition on unit sludge handling processes is provided under those sections of this chapter which deal with sludges derived from use of specific chemicals.

5.2.2 Points of Chemical Addition and Methods of Combining Sludges

Table 5-1 shows a breakdown of the 174 plants by point(s) of chemical addition and by how the chemical sludges are combined with organic sludges at different locations within the plant.

Chemical addition to secondary treatment was the most common option practiced, accounting for 62 percent of the plants surveyed. Twenty-six percent of the plants added chemicals to primary treatment, 5 percent of the plants added chemicals to both primary and secondary treatment, and 6 percent of the plants added chemicals to tertiary facilities.

Of the plants practicing chemical addition to primary treatment and generating both primary and secondary sludges, 94 percent of the plants combined their

Table 5-1. Combination of Chemical Sludges with Other Sludges for Processing as Practiced by Plants in EPA Survey (1).

Point at Which Chemical Sludges and Other Sludges were Combined	Number of Plants by Point(s) of Chemical Addition						Total
	Primary Addition	Secondary Addition	Tertiary Addition	Primary and Sec. Addition	Primary and Tert. Addition	Sec. and Tert. Addition	
<i>Plants w/o primary or secondary clarification</i>							
No primary clarifier	NA ¹	28	1	NA	NA	1	30
No secondary clarifier	13	NA	NA	NA	NA	NA	13
<i>Plants with primary and secondary clarification</i>							
Sludges not combined	2	10	6	1	0	0	19
Combined in or before primary clarifier	12	26	0	3	0	0	41
Combined in or before thickener	13	14	2	1	1	0	31
Combined in or before digester	6	24	2	3	0	0	35
Combined in or before dewatering device	0	6	0	0	0	0	6
Total	46	108	11	8	1	1	175²

¹ Not applicable.

² One plant with tertiary addition is counted twice because it has no primary sludge and it has secondary and tertiary sludges which are not combined.

chemical primary sludges with organic secondary sludge before processing. Of these plants combining their sludges, 39 percent accomplished this by returning secondary sludge to the primary clarifiers. The remainder combined their sludges in or before thickening, stabilization, or dewatering processes.

Of the plants practicing chemical addition to secondary treatment and which generated both primary and secondary sludges, 88 percent combined their sludges before processing. Of the plants combining their sludges, 37 percent practiced this by returning secondary sludge to the primary clarifier.

5.2.3 Sludge Generation Rates and Characteristics

The additional sludge generated from use of chemicals for phosphorus precipitation is a key design consideration. Although chemical addition to the liquid stream is easily implemented, the impacts on thickening, stabilization, dewatering, and disposal operations are often severe. This is due to both the additional mass and volume of sludge generated as well as the effects on thickening and dewatering characteristics.

Figure 5-1 summarizes volumes of sludge produced (before thickening) per cubic meter of influent flow as reported in the EPA survey of 174 plants (1). It is apparent that lime addition produced significantly greater volumes of sludge than mineral salt addition. Chemical addition to secondary processes produced greater volumes of sludge than chemical addition to the primary when considering primary and secondary processes separately. However, on a total plant basis, the additional sludge volumes resulting from chemical

addition were approximately equivalent for both primary and secondary chemical addition.

Average increases in sludge volume before thickening were 25 percent for iron salt addition, 58 percent for aluminum salt addition, and several hundred percent for lime addition (1). Bear in mind that these figures are volumes before thickening, and as such do not take into account the relative thickening characteristics of sludge which will affect volumes requiring dewatering; nor do they consider dewatering characteristics which will affect volumes requiring final disposal.

It is important to note that the reported sludge volumes represent a diverse array of plants with wide ranges of wastewater characteristics, treatment process configurations, operating modes, and effluent quality, all of which significantly impact sludge generation rates. Therefore, these figures are for illustrative purposes only, and should not be used for design.

Figure 5-2 summarizes mass sludge production data from the EPA survey (1). Values are expressed as kilograms of dry solids per cubic meter of influent flow. Again, it should be noted that these figures represent plants with wide ranges of wastewater and process characteristics, which can significantly impact the mass of sludge generated. Use of these figures for design purposes or chemical selection is not recommended.

Table 5-2 summarizes data on solids content of chemical sludges at the 174 plants surveyed in the EPA-sponsored study. The total and volatile solids content are for the sludges prior to thickening or other

Figure 5-1 Volumetric sludge production from EPA survey - m³ sludge/m³ plant influent (1).

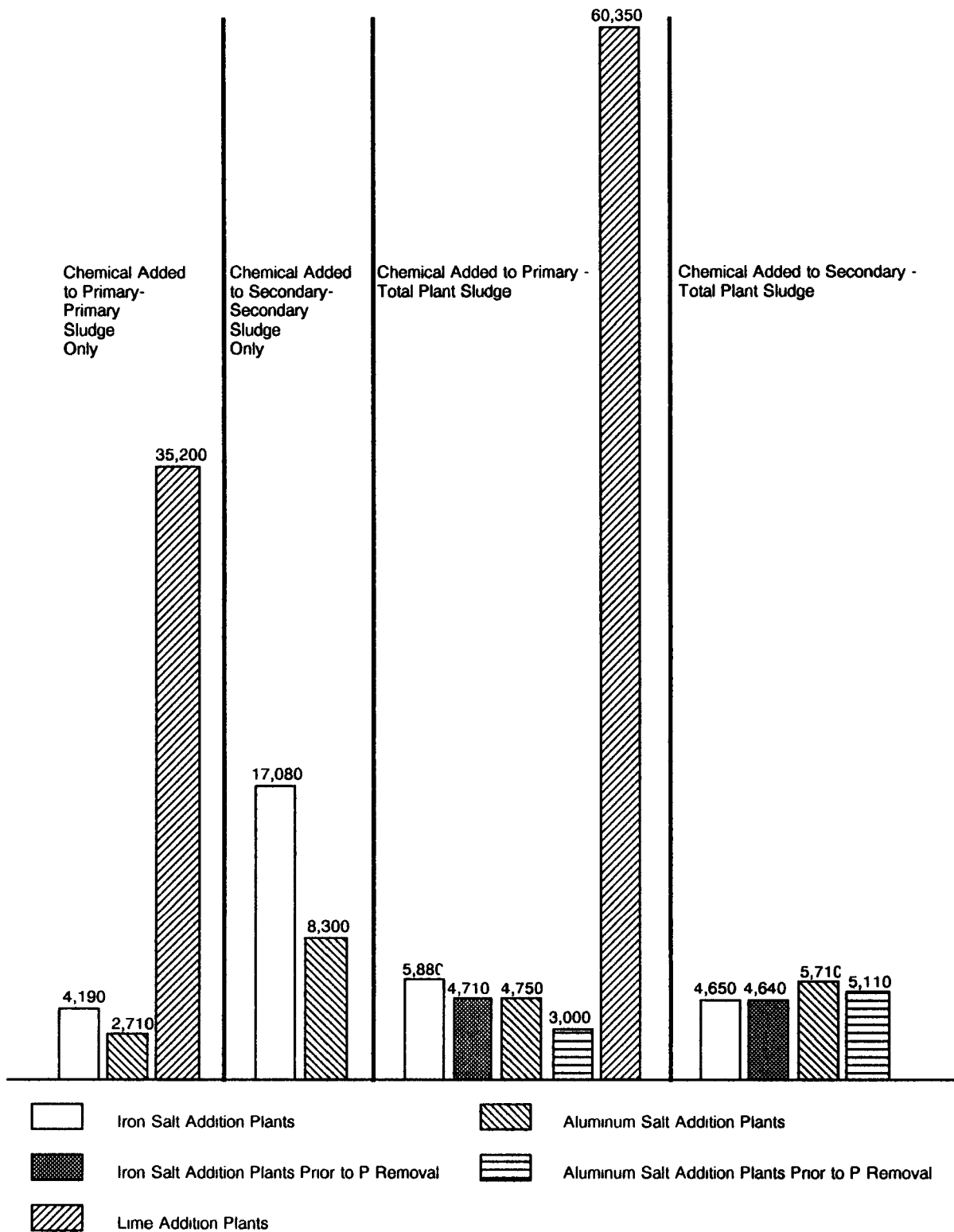
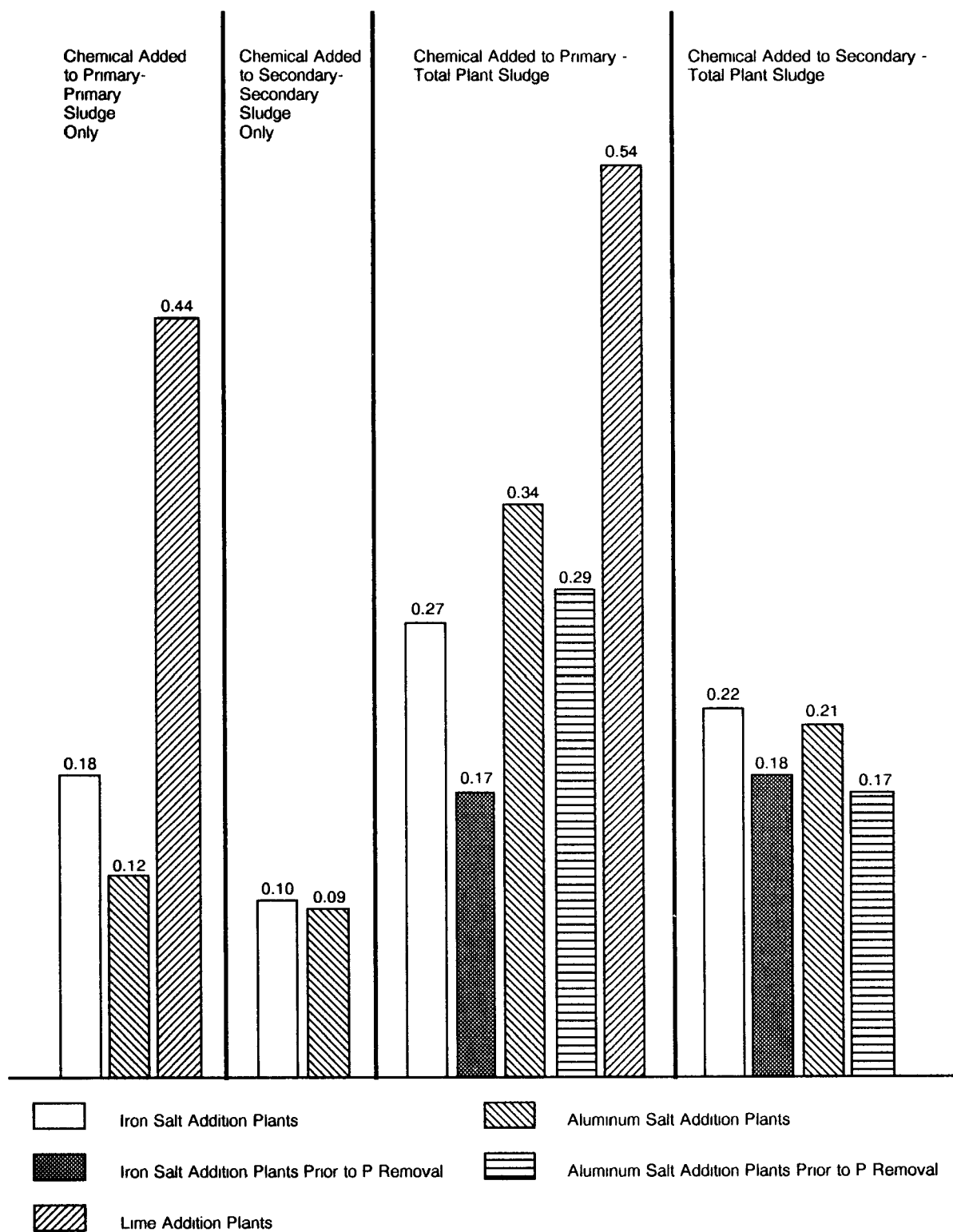


Figure 5-2 Mass sludge production from EPA survey - kg sludge/m³ plant influent (1).



processing. For chemical addition to primary treatment, iron sludges had the highest average total solids content at 5.26 percent compared to alum sludge at 3.95 percent and lime sludge at 1.1 percent. For chemical addition to secondary treatment, secondary biological-iron sludge had an average total solids content of 0.93 percent vs. 1.41 percent for biological-aluminum sludge. When combined with primary sludges, the sludge solids contents for iron and aluminum addition to secondary treatment were similar at 4.13 and 3.82 percent, respectively. Although these data are useful for illustrative purposes, the values are averages obtained from plants with different wastewater and process characteristics. This may explain some inconsistencies in the data. For example, for iron addition to primary treatment, total solids for primary sludge is 5.26 percent, while total solids for the combined (total plant) sludge is 5.73 percent. In practice, this would not occur, since secondary sludge always has a significantly lower total solids content than primary sludge.

5.2.4 Prevalence of Various Sludge Treatment and Disposal Options

Table 5-3 summarizes the responses from plants contacted in the EPA survey regarding the types of processes employed for thickening, stabilization, conditioning, dewatering, drying, reduction, and disposal of sludge.

Of the thickening options, gravity thickening was by far the most prevalent, as would be expected. In most cases, flotation thickening was used for secondary sludge only. The five plants which used centrifuges for thickening added lime to primary treatment for phosphorus removal.

Aerobic and anaerobic digestion were the most prevalent stabilization techniques. Of the plants using iron salts, 20 percent used aerobic digestion. Of the plants using aluminum salts, 46 percent used aerobic digestion, while 54 percent used anaerobic digestion.

Chemical conditioning was the most prevalent method of conditioning, followed by thermal conditioning. Seventy-five percent of the plants using iron salts and 74 percent of the plants using aluminum salts employed chemical conditioning.

Sand drying beds and vacuum filters were the most prevalent dewatering techniques. Of the plants using iron salts, 49 percent used sand drying beds and 34 percent used vacuum filters. Of the plants using aluminum salts, 44 percent used sand drying beds, 24 percent used vacuum filters, and 11 percent used centrifuges. None of the plants using iron employed centrifuges for dewatering.

Three plants in the survey reported using heat dryers for further sludge volume reduction, and 22 plants employed multiple hearth or fluidized bed incineration.

The most prevalent means of final sludge disposal was application to agricultural land, lawns, or gardens, accounting for 52 percent of the plants surveyed. This was followed by sanitary landfill (27 percent), private or authority-owned dump site (16 percent), and land reclamation (5 percent).

5.3 Sludge Derived from Addition of Aluminum Salts

5.3.1 Sludge Characteristics

Aluminum salts may be employed for phosphorus removal by addition to primary, secondary, or tertiary treatment processes. Secondary addition is the most common, accounting for approximately 82 percent of the plants using aluminum salts as determined by the EPA survey (1). Addition of aluminum to primary treatment accounted for only about 14 percent of the plants using aluminum salts. The few remaining plants added aluminum salts to tertiary treatment (3 percent) or to both primary and secondary treatment (1 percent).

Sludge characteristics are dependent on numerous factors including wastewater composition, chemical dosage, point of chemical addition (primary, secondary or tertiary), whether the various sludges are combined before processing and in what proportions, detention time in clarifiers or holding tanks, and other factors. Although it is difficult to predict the impact of chemical addition on thickening and dewatering characteristics, two conclusions can be drawn with some certainty:

1. That the addition of aluminum salts will result in an increase in sludge volume and mass, and
2. The thickening and dewatering characteristics will likely be different than if no chemical were added.

Actual experience with processing chemical - biological sludges has been extremely variable. In some cases, chemical addition has improved thickening and dewatering characteristics, while in other cases chemical addition has had a detrimental impact. Many problems can be traced to overloading of unit processes due to increased volumes of sludge from chemical addition.

Knowledge of general sludge characteristics and chemical dosages is generally of little value in predicting the amenability of sludges to thickening, dewatering, and stabilization processes. This is due to the interaction and impact of the host of factors which affect performance of sludge processing operations. Design of sludge handling systems must therefore be

Table 5-2. Combination of Chemical Sludges with Other Sludges for Processing as Practiced by Plants in EPA Survey (1).

Type of Chemical Sludge and Whether Combined with Other Plant Sludge(s)	Sludge Characteristics (before processing)			
	Total Solids, percent		Volatile Solids, percent of TS	
	Range	Average	Range	Average
<u>Iron addition to primary</u>				
Primary sludge	3.4 - 8.0	5.26	45 - 69	57
Total plant sludge	2.31 - 10.0	5.73	40 - 70	57
<u>Iron addition to secondary</u>				
Secondary sludge	0.2 - 4.0	0.93	50 - 70	62
Total plant sludge	0.5 - 7.75	4.13	45 - 72	62
<u>Iron addition to tertiary</u>				
Tertiary sludge	4.0	4.0	35	35
Total plant sludge	4.64 - 5.0	4.82	62	62
<u>Alum addition to primary</u>				
Primary sludge	3.3 - 43.5	3.95	61 - 67	65
Total plant sludge	3.96 - 5.0	4.49	46 - 70	59
<u>Alum addition to secondary</u>				
Secondary sludge	0.4 - 4.4	1.41	60 - 78	67
Total plant sludge	1.0 - 7.0	3.82	52 - 70	59
<u>Lime addition to primary</u>				
Primary sludge	0.7 - 1.5	1.1	N/A ¹	N/A
Total plant sludge	0.64 - 0.82	0.73	N/A	N/A
<u>Lime addition to tertiary</u>				
Tertiary sludge	2.5 - 4.0	3.3	11 - 30	21
Total plant sludge	1.95	1.95	39	39

¹ Not available.

based, at a minimum, on laboratory tests such as settlometer tests, specific resistance, filterleaf, capillary suction time, and other tests. Where possible, pilot- or full-scale units should be used to establish design criteria for full-scale sludge thickening and dewatering devices. Specific tests used to characterize sludges resulting from chemical phosphorus precipitation processes are discussed in subsequent sections of this manual.

5.3.2 Sludge Generation Rates

As previously discussed, addition of chemicals to remove phosphorus will increase the mass of sludge generated. This increase can be attributed to three components (2):

1. Formation and removal of chemical solids such as metal phosphates and metal hydroxides,
2. Improved removals of organic solids during clarification, and
3. Removal of dissolved solids.

Several methods are available to estimate sludge generation rates for design purposes. When an existing plant is to be upgraded for chemical phosphorus removal, the preferred approach is to conduct full-scale trials at the facility and measure sludge production under expected operating conditions. This allows optimization of chemical dosages and determination of the impact of

operational changes such as modification of sludge age. For new facilities, pilot plant tests are the best approach to estimate sludge generation rates. However, such studies are costly and may not be economically feasible for small plants. In such cases, jar tests should be performed to provide data to estimate additional sludge generation resulting from chemical addition, although such tests may not be representative of full-scale, dynamic conditions. Procedures are also available to allow calculation of the theoretical mass of solids generated from wastewater treatment schemes employing chemical phosphorus precipitation (2-4). An approach is described below for estimating the quantity of additional sludge resulting from use of alum for phosphorus removal. A procedure for calculating baseline sludge production (no chemical addition) has not been shown, since this is described in detail in Reference 4 for several types of treatment (e.g. processes activated sludge, trickling filter, etc.).

Additional sludge generated as a result of alum addition is due to formation and removal of chemical solids, improved removal of suspended solids, and removal of dissolved solids. Another consideration in estimating sludge production is that the quantity of biological sludge which is generated may be affected by selection of the point of chemical addition. For example, while alum addition to the primary clarifier will increase the production of primary sludge, improved BOD and SS removal in primary treatment will reduce the organic loading to downstream

Table 5-3. Prevalence of Treatment and Disposal Processes for Chemical Sludges Among Plants in EPA Survey (1)*.

Sludge Treatment and Disposal Unit	Major Chemical Used and Point of Addition									Total
	Iron Salt			Aluminum Salt			Lime			
	Primary	Sec.	Tertiary	Primary	Sec.	Tertiary	Primary	Sec.	Tertiary	
<u>Thickening</u>										
Gravity	18	25	2	4	18	1	8	0	0	76
Flotation	1	4	0	0	5	0	0	0	0	10
Centrifuge	0	0	0	0	0	0	5	0	0	5
<u>Stabilization</u>										
Composting	1	0	0	1	0	0	0	0	0	2
Aerobic digestion	4	11	1	1	24	0	0	0	0	41
Anaerobic digestion	27	37	1	5	25	1	1	0	0	97
Lime stabilization	0	0	0	1	0	0	0	0	0	1
<u>Conditioning</u>										
Chemical conditioning	11	12	1	2	12	0	2	0	0	40
Elutriation	1	2	0	0	0	0	0	0	0	3
Thermal conditioning	2	4	0	1	4	0	0	0	0	11
<u>Dewatering</u>										
Pressure filter	1	3	0	1	1	0	2	0	1	8
Sand drying bed	17	20	0	2	18	0	1	0	0	58
Centrifuge	0	0	0	1	4	0	0	0	1	6
Vacuum filter	12	13	1	1	10	0	0	0	0	37
Lagoon	4	4	1	1	3	1	0	0	2	16
Hor. moving screen concentrator	0	0	0	0	2	0	0	0	1	3
Cylindrical rotating gravity filter	0	0	0	0	1	0	0	0	0	1
<u>Heat Drying</u>										
Flash dryer	0	0	0	0	1	0	0	0	0	1
Multiple hearth dryer	0	1	0	0	1	0	0	0	0	2
<u>Reduction</u>										
Incineration	7	5	0	0	5	0	5	0	0	22
<u>Final Disposal</u>										
Land reclamation	0	1	0	1	6	0	1	0	0	9
Sanitary landfill	16	18	1	1	13	0	4	0	0	53
Agricultural fields lawns, gardens	18	39	1	7	30	1	5	0	1	102
Private- or authority- owned dump site	8	10	0	2	8	0	2	0	1	31

* One plant may use more than one method of thickening, dewatering, etc.

biological processes and, hence, reduce the quantity of biological solids generated. A procedure for estimating quantities of additional sludge resulting from alum addition is given below.

a. Chemical sludge

This calculation assumes that the aluminum reacts with phosphorus compounds first, and that excess aluminum forms aluminum hydroxide.

Given:

$$P_{in} = 8 \text{ mg/l}$$

$$P_{out} = 1 \text{ mg/l}$$

$$\text{Al:P dose} = 2.2:1$$

$$\text{Atomic weight of P} = 31$$

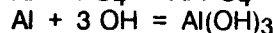
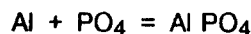
$$\text{Atomic weight of Al} = 27$$

$$\text{Atomic weight of AlPO}_4 = 122$$

$$\text{Atomic weight of Al(OH)}_3 = 78$$

$$\text{Al dose} = 2.2 \times 8 \times (27/31) = 15 \text{ mg/l}$$

Stoichiometry:



$$(8 - 1 \text{ mg/l P}) \div 31 = 0.23 \text{ mmole/l AlPO}_4 \text{ produced}$$

$$(15 \text{ mg/l Al}) \div 27 = 0.56 \text{ mmole/l Al added}$$

$$0.56 - 0.23 = 0.33 \text{ mmole/l Al in excess to Al(OH)}_3$$

$$\text{AlPO}_4 \text{ sludge: } 0.23 \text{ mmole/l} \times 122 = 28.1 \text{ mg/l AlPO}_4$$

$$\text{Al(OH)}_3 \text{ sludge: } 0.33 \text{ mmole/l} \times 78 = 25.7 \text{ mg/l Al(OH)}_3$$

$$\text{Total chemical sludge: } 28.1 \text{ mg/l AlPO}_4 \\ 25.7 \text{ mg/l Al(OH)}_3$$

53.8 mg/l Al sludge

Chemical sludge produced = 53.8 mg sludge/liter of wastewater treated (449 lb/10⁶ gallons).

Because the stoichiometry is only an approximation of the chemical reactions which occur, and because some data indicate greater quantities of chemical sludge than predicted, it has been recommended that the calculated sludge production value be increased by 35 percent (2).

Design estimate of chemical sludge production:

$$1.35 \times 53.8 \text{ mg/l} = 72.6 \text{ mg sludge/liter of wastewater treated (605 lb/10}^6 \text{ gallons)}$$

b. Sludge from improved removal of suspended solids

Additional sludge resulting from improved suspended solids removal is calculated by assuming a greater removal efficiency. For example, in primary clarification without chemical addition, removals of suspended solids are typically 50 percent. With alum addition, however, removal efficiencies may be 75 percent or greater. The additional sludge generated by chemical addition to the primaries can be calculated as shown below.

Given:

SS_{in} = 200 mg/l
SS removal efficiency (no chemicals) = 50 percent
SS removal efficiency (with alum) = 75 percent

Additional sludge generated:

$$(0.75 - 0.50) \times 200 \text{ mg/l} = 50 \text{ mg sludge/liter of wastewater treated (417 lb/10}^6 \text{ gallons)}$$

When aluminum is added to the secondary biological process such as the activated sludge basin, additional sludge production from improved SS removal may not be evident. However, generation of chemical solids (AlPO₄ and Al(OH)₃) can be assumed to be approximately equal regardless of the point of chemical addition. For a well operated standard rate activated sludge plant, addition of aluminum salts to the aeration basin may have little impact on clarification efficiency. However, for some high rate activated sludge or trickling filter plants, secondary clarification efficiency may improve significantly, resulting in additional sludge production. An improvement in effluent suspended solids concentration from 20 to 10 mg/l will result in 10 mg of additional sludge per liter of wastewater treated (83 lb per million gallons).

For tertiary applications of aluminum following secondary biological treatment, additional sludge

resulting from improved removal of SS will generally be small. Estimates of tertiary sludge production from SS removal are based on anticipated SS concentrations in secondary and tertiary effluents.

c. Sludge from removal of dissolved solids

Data exist to show that dissolved solids are removed as a result of chemical addition (2). Reported removals of soluble TOC are about 30% percent using alum. Removal of soluble COD has been reported to be approximately 40 percent. The sludge mass resulting from removal of dissolved solids must be estimated indirectly from soluble TOC, COD, or BOD loadings. The following relationships can be used for this purpose. The derivation of these relationships may be found in Reference 2.

Sludge mass resulting from removal of dissolved solids (assuming 30 percent removal of soluble organics)

$$\begin{aligned} &= \text{STOC}_{\text{in}} (\text{mg/l}) \times 0.30 \times 2.5 \times 1.18 \\ &= \text{SCOD}_{\text{in}} (\text{mg/l}) \times 0.30 \times 1.1 \times 1.18 \\ &= \text{SBOD}_{\text{in}} (\text{mg/l}) \times 0.30 \times 1.6 \times 1.18 \end{aligned}$$

where:

STOC_{in} = soluble TOC in influent
SCOD_{in} = soluble COD in influent
SBOD_{in} = soluble BOD in influent

The latter two equations are only applicable to influents prior to biological oxidation processes (2).

d. Design example

Estimate the sludge production from a 3,785-m³/d (1-mgd) conventional activated sludge plant with and without alum addition to the primary clarifier.

Given:

P_{in} = 8 mg/l
BOD_{in} = 200 mg/l
SS_{in} = 220 mg/l
VSS = 0.75 x SS
STOC_{in} = 50 mg/l

Sludge from primary treatment without alum addition:

Assume 50 percent SS removal, 30 percent BOD removal;

$$0.5 \times 220 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001 = 416 \text{ kg/d (917 lb/d)}$$

Sludge from primary treatment with alum addition:

Assume 90 percent P removal, 75 percent SS removal, 50 percent BOD removal, 30 percent STOC

removal (from test results); assume Al:P dosage ratio of 2.0 (from test results)

Al dose = $(2.0 \times 8 \text{ mg/l}) \times (27/31) = 13.9 \text{ mg/l Al}$
P removed = $0.90 \times 8 \text{ mg/l} = 7.2 \text{ mg/l P}$
 AlPO_4 sludge = $7.2/31 = 0.23 \text{ mmole/l AlPO}_4$
Total Al added = $13.9/27 = 0.51 \text{ mmole/l Al}$
Excess Al = $0.51 - 0.23 = 0.28 \text{ mmole/l Al}$
 AlPO_4 sludge = $0.23 \times 122 = 28.1 \text{ mg/l AlPO}_4$
 Al(OH)_3 sludge = $0.28 \times 78 = 21.8 \text{ mg/l Al(OH)}_3$

Total chemical sludge produced:
 $28.1 \text{ mg/l} + 21.8 \text{ mg/l} = 49.9 \text{ mg/l}$

Design chemical sludge production:
 $49.9 \text{ mg/l} \times 1.35 = 67.4 \text{ mg/l}$

or

$67.4 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001 = 255 \text{ kg/d} (562 \text{ lb/d})$

Sludge from SS removal:
 $0.75 \times 220 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001 = 625 \text{ kg/d}$
 $(1,375 \text{ lb/d})$

Sludge from removal of dissolved solids (DS):
 $50 \text{ mg/l} \times 0.30 \times 2.5 \times 1.18 = 44.2 \text{ mg/l}$

or

$44.2 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001 = 167 \text{ kg/d} (368 \text{ lb/d})$

Secondary sludge - no alum addition to primary (from reference 4):

$$\text{WAS}_T = P_x + I_{nv} - E_T$$

$$P_x = Y (S_r) - K_d (M)$$

$$= (Y) (S_r) / (1 + K_d \text{ SRT})$$

where:

WAS_T = waste activated sludge production, kg/d
 P_x = net growth of biological solids (VSS), kg/d
 I_{nv} = inert SS fed to process, kg/d
 E_T = effluent SS, kg/d
 Y = gross yield coefficient, kg/kg
 S_r = BOD removed, kg/d
 K_d = decay coefficient, day⁻¹
 M = system inventory of microbial solids (VSS), kg
SRT = solids retention time, days

Assume:

$Y = 0.67 \text{ kg/kg}$ from test data
 $K_d = 0.06 \text{ day}^{-1}$ from test data
 $\text{BOD}_{out} = 15 \text{ mg/l}$
SRT = 10 days

$\text{BOD}_{in} - \text{BOD}_{out} = (1 - 0.3) (200 \text{ mg/l}) - 15 \text{ mg/l}$

$$S_r = 125 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001$$

$$= 473 \text{ kg/d}$$

$$P_x = (0.67 \text{ kg/kg} \times 473 \text{ kg/d}) / [1 + (.06 \text{ d}^{-1}) (10 \text{ d})]$$

$$= 198 \text{ kg/d}$$

$$\text{Inert SS feed} = (1 - 0.5) (1 - 0.75) (220 \text{ mg/l})$$

$$= 27.5 \text{ mg/l}$$

$$I_{nv} = 27.5 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001$$

$$= 104 \text{ kg/d}$$

Assume effluent SS = 20 mg/l;

$$E_T = 20 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001$$

$$= 76 \text{ kg/d}$$

$$\text{WAS}_T = P_x + I_{nv} - E_T$$

$$= 198 \text{ kg/d} + 104 \text{ kg/d} - 76 \text{ kg/d}$$

$$= 226 \text{ kg/d} (498 \text{ lb/d})$$

Secondary sludge - alum addition to primary:

Calculate reduced BOD loading due to improved BOD removal in primaries;

$$\text{BOD}_{in} = (1 - 0.5) (200 \text{ mg/l}) (3,785 \text{ m}^3/\text{d}) (0.001)$$

$$= 379 \text{ kg/d}$$

$$\text{BOD}_{out} = 15 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001$$

$$= 57 \text{ kg/d}$$

(assumes no improvement in secondary effluent quality)

$$S_r = 379 \text{ kg/d} - 57 \text{ kg/d} = 322 \text{ kg/d}$$

$$P_x = (0.67 \text{ kg/kg} \times 322 \text{ kg/d}) / [1 + (.06 \text{ d}^{-1}) (10 \text{ d})]$$

$$= 135 \text{ kg/d}$$

$$\text{Inert SS feed} = (1 - 0.75) (1 - 0.75) (220 \text{ mg/l})$$

$$= 13.8 \text{ mg/l}$$

$$I_{nv} = 13.8 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001$$

$$= 52 \text{ kg/d}$$

Assume effluent SS = 20 mg/l;

$$E_T = 20 \text{ mg/l} \times 3,785 \text{ m}^3/\text{d} \times 0.001$$

$$= 76 \text{ kg/d}$$

(assumes no improvement in secondary effluent quality)

$$\text{WAS}_T = P_x + I_{nv} - E_T$$

$$= 135 \text{ kg/d} + 52 \text{ kg/d} - 76 \text{ kg/d}$$

$$= 111 \text{ kg/d} (224 \text{ lb/d})$$

Summary of sludge production calculations:

	Sludge production, kg/d	
	w/o alum	w/alum
<u>Primary clarifier</u>		
SS sludge	416	625
DS sludge	0	167
Chemical Sludge	0	255
Total	416	1,047
<u>Secondary clarifier</u>		
WAS	226	111
<u>Total sludge</u>	642	1,158

5.3.3 Thickening

5.3.3.1 Gravity Thickening

Relatively little definitive design and performance information is available in the literature for gravity thickening of alum sludges. Thickening characteristics of sludges generated from plants practicing chemical phosphorus precipitation will vary depending on wastewater characteristics, point of chemical addition, chemical dosage, wastewater treatment processes employed, whether sludges are combined before thickening, relative proportions of chemical and biological sludges, sludge characteristics before thickening, and other factors. Because of the wide variability in thickening characteristics, thickening tests should be conducted whenever the actual sludge is available, as in the case of retrofitting or expanding existing facilities. In such cases, full-scale trials with chemical addition are recommended to generate representative samples of sludge for conducting thickening tests. Procedures for conducting such tests may be found elsewhere (5-8).

In some cases where new treatment facilities are to be designed, the design engineer does not have the benefit of having existing sludges on which to conduct thickening tests, and must rely on published design guidelines. Table 5-4 provides design criteria for thickeners receiving various types and combinations of alum and non-chemical sludges.

In addition to mass loading criteria shown in Table 5-4, hydraulic loading must also be considered in the design of gravity thickeners. For primary sludges, typical maximum overflow rates are 24-30 m³/m²/d (600-720 gpd/ft²). However, for waste activated sludge or combinations of primary and waste activated sludges, hydraulic loadings rates should be lower than this, 4-8 m³/m²/d (100-200 gpd/ft²) (4).

Experience with thickening of alum sludges in combination with non-chemical sludges has been highly variable. Some reports indicate an improvement in thickening characteristics of sludges

Table 5-4. Typical Gravity Thickener Design Criteria.

Sludge Type	Influent Solids Conc. percent	Expected Underflow Conc. percent	Mass Loading kg/m ² /hr	Ref.
Primary	2-7	5-10	3.9-5.9	9
Primary w/alum ¹	2.1-3.7	2.5-6.7	0.4-1.0	10
Primary + WAS	0.5-1.5	4-6	1.0-2.9	8
(Primary w/alum) + WAS	-	-	-	-
Primary + (WAS w/alum)	0.2-0.4	4.6-6.5	2.4-3.4	8
WAS	0.5-1.5	2-3	0.5-1.5	9
WAS w/alum	0.5-1.5	1.5-2.0	0.6-0.8	11
Tertiary w/alum ²	0.5-1.5	2.5-3.0	0.6-1.0	12

¹ Data reflect use of anionic polymer to assist in thickening.

² From water treatment plant sludge containing alum and powdered activated carbon.

when alum addition was initiated; others show detrimental impacts on thickener performance (1,3). Unfortunately, in the latter cases, it is difficult to determine whether poor performance was due to the change in sludge characteristics or the additional solids loading from increased sludge production.

In pilot-studies on alum-primary sludges, it was found that underflow solids concentration generally decreased as the amount of chemical solids in the feed sludge increased (10). Note that in Table 5-4 recommended mass loadings to gravity thickeners decrease substantially for primary sludge with alum vs. primary sludge alone.

Several investigators have noted an improvement in the settlability (decrease in SVI) of mixed liquor solids where aluminum salts are dosed directly into the aeration basin (14-16). Little data are available regarding thickening characteristics of the resulting sludges, however. In general, where alum is added to the aeration basin of the activated sludge process, recommended loadings to gravity thickeners are similar to those for which no chemicals are added.

No data were found for thickening of tertiary wastewater sludges resulting from alum addition. In many cases, alum is added to the secondary effluent immediately prior to tertiary filtration without intermediate clarification. Tertiary alum sludges would be expected to behave similarly to alum sludges from potable water treatment.

Addition of polymer to the influent sludge of a gravity thickener has been found to improve solids capture and increase underflow solids concentrations (1, 10).

5.3.3.2 Flotation Thickening

Flotation thickening is often employed for thickening of waste activated sludge from extended aeration

plants without primary clarification, or for secondary sludges where primary and secondary sludges are thickened separately. In the majority of cases, flotation thickening is not applied to primary sludges or combined primary-secondary sludges. Where chemicals are added to the primary clarifier for phosphorus removal, organic loadings to downstream biological processes are reduced substantially, thus reducing the quantity of biological sludge generated. In such cases, separate thickening of primary and secondary sludges is not justified and the sludges would most likely be combined prior to thickening by gravity.

In the EPA-sponsored survey of plants removing phosphorus, it was found that 6 percent of the facilities employed flotation thickening (1). In all but one of these plants, flotation thickening was used on iron or aluminum waste activated sludge alone. Half of these facilities aerobically digested the sludge before thickening. Of the nine plants surveyed, four reported significant changes in flotation thickener performance when processing chemical sludges; four reported no change at all. The ninth plant did not comment on the impacts of chemical sludge handling. Of the four plants reporting significant impacts, three reported negative impacts, which included:

1. The need to reduce hydraulic loading rate,
2. The need to use both anionic and cationic polymers, and
3. A decrease in float (thickened) solids concentration.

The plant reporting a positive impact indicated an increase in float solids concentration. Some plants noted improved performance with polymer addition.

Of the nine plants that used flotation thickening, four added alum to the aeration basins, and two of these plants aerobically digested the sludge prior to thickening. Feed solids concentrations were 0.5-1.5 percent, while float solids were typically 3.5-5 percent (1).

In batch laboratory flotation tests on aerobically digested waste activated sludge from a contact-stabilization plant, no difference was found between the thickening properties of the control digester and alum digester sludge. The sludge was concentrated from 0.75 percent TSS to 3.5 percent TSS without polymer addition at an air: solids ratio of 0.03 (17).

Mass loadings of waste activated sludge (no chemicals) to dissolved air flotation thickeners vary widely in practice, ranging from 2.8 to 34.0 kg/m²/hr (0.6-6.9 lb/ft²/hr) (4). Typical design loadings for flotation thickeners receiving non-chemical waste activated sludge are 2.4-3.9 kg/m²/hr (0.5-0.8

lb/ft²/hr) without polymer and 4.9-7.3 kg/m²/hr (1.0-1.5 lb/ft²/hr) with polymer (8) [In Chapter 4 of Reference 8, Sludge Thickening, WPCF MOP No. FD-1, all solids loadings shown in metric units are calculated incorrectly; use only values shown in English units; kg/m²/hr = 4.9 x lb/ft²/hr].

For dissolved air flotation thickening of waste activated sludge with alum, solids loading rates of 10-24 kg/m²/hr (2-5 lb/ft²/hr) with polymer have been recommended (11).

Pilot-studies have been conducted on flotation thickening of chemical primary sludges alone (10). Unfortunately, flotation thickening is seldom used for primary sludge. Solids loadings for alum sludges were 6.8-24.9 kg/m²/hr (1.4-5.1 lb/ft²/hr). Feed solids (TSS) were 0.8-2.7 percent, and float solids (TSS) were typically 2.5-5.0 percent. Polymer addition was necessary to achieve good performance (10).

5.3.4 Stabilization

5.3.4.1 Aerobic Digestion

In the EPA survey of 174 plants removing phosphorus, 41 plants (24 percent) employed aerobic sludge digestion (1). Only four plants reported problems with stabilization of chemical sludge; the majority of these problems were directly attributed to increased sludge volumes from chemical addition. In some cases, generation of a more concentrated sludge as a result of alum addition necessitated increasing the air supply to maintain adequate mixing of the digester contents.

No deterioration of supernatant quality was reported as a result of increased sludge volumes or changed sludge characteristics upon chemical addition (1).

Laboratory studies on waste activated sludge with alum showed little impact of the presence of aluminum precipitates on the aerobic digestion process (18,19).

In a full-scale study of alum addition at a 1.7 mgd contact-stabilization plant, TSS reduction in that alum sludge during aerobic digestion was only 12 percent vs. 31 percent in the control sludge. This does not provide an adequate measure of the degree of organics oxidation which occurred, however, since the alum sludge would be expected to have a higher percentage of non-volatile solids. Data on volatile solids reduction were not provided. The digested alum sludge thickened significantly better than the control when subjected to laboratory tests (17).

There has been some concern regarding the possible release of phosphorus from the solid to the liquid phase during anoxic storage of aerobically digested phosphorus-laden sludges. Laboratory studies were conducted on untreated primary sludge and mixed

primary-chemical sludges to address this issue (20). It was found that during storage of primary sludge at different degrees of stability (digestion times), phosphorus was always released from the solid phase to the liquid phase. Orthophosphorus concentration increased from an initial 5-10 mg/l P to 25-50 mg/l P after 10-12 days of anoxic storage. However, for mixed primary and alum secondary sludge, no release of orthophosphate took place during anoxic storage, regardless of the degree of stability and digester temperature (20).

5.3.4.2 Anaerobic Digestion

Review of the literature on anaerobic digestion of alum sludges indicates mixed results regarding the impact of aluminum addition on the digestion process. In one laboratory study, it was found that alum-precipitated phosphorus concentrated in the sludge of the digester and was not released during anaerobic digestion. Phosphorus concentrations in the supernatant from the digester receiving alum sludge were less than in the control. No toxic effects of the alum sludge were noted, as evidenced by volatile acid and gas production characteristics (21). Similar results were found in another laboratory study (22).

However, in a closely controlled laboratory study, it was found that chemical coagulation of organic materials with alum caused a significant decrease in the anaerobic digestibility of the resulting sludge as measured by gas production per mass of organic solids in the feed. This was attributed to "association of substrate with coagulant floc, rendering a portion of the organics less accessible and/or reactive to microorganisms or their extracellular enzymes" (23).

Other laboratory-scale studies showed digester performance, as measured by gas production, methane production, volatile solids reduction, and COD reduction, decreased with increasing alum dosages. At an alum dosage of 200 mg/l, digester performance was 92 percent of the control value, decreasing to 82 percent at an alum dosage of 400 mg/l. Reduced alkalinity in the digesters receiving alum sludge was noted (24). At the plant from which the sludges were derived for the above laboratory tests, gas production decreased to about 50 percent of normal levels upon addition of 250 mg/l of alum to the raw wastewater (24).

In the EPA survey of plants removing phosphorus, 56 percent of plants employed anaerobic digestion. Of these, 22 percent reported that chemical addition was having a significant impact on their digestion process. Negative impacts reported included (1):

1. Increased energy requirements for sludge mixing, pumping, and heating,
2. Difficulty in achieving adequate digester mixing and heating,

3. Increased labor requirements for sludge pumping,
4. Poor solids-liquid separation, and
5. Reduction in digester efficiency.

At Richardson, Texas, solids stratification and digester upset occurred when alum was added to the raw wastewater. When the point of alum addition was moved to the secondary process, no digestion problems were observed. At Ashland, Wisconsin, alum was added to the secondary treatment stage (step aeration activated sludge), and sludge was stabilized with two stage digestion. After commencement of alum addition, solids-liquid separation no longer occurred. Polymer addition did not solve the problem. Similar problems in solids-liquid separation occurred in the secondary digesters at Three Rivers, Michigan and Gladstone, Michigan. Several plants reported significant increases in sludge volume which reduced detention times and exceeded heat exchanger capacity (1).

Special considerations should be given to the design of anaerobic digesters receiving chemical sludges. The most important is the capability to handle increased sludge volumes resulting from chemical addition. Procedures for estimating sludge mass have been provided in this chapter. Sludge volume estimates require knowledge of sludge concentrations resulting from clarification and thickening, which can be best determined from pilot or full-scale trials. In existing plants, expansion of thickener capacity and/or addition of polymer to thickeners can reduce the volume of sludge fed to the digesters. In addition to sludge volume increases, consideration should be given to potential performance inhibition from alum addition, which may justify longer detention times.

Mixing is also very important in order to achieve optimal volatile solids destruction and gas production. Mixing maintains contact between the active biomass and the substrate; creates physical, chemical, and biological uniformity throughout the digester; disperses metabolic end products and toxic materials; and prevents formation of scum layer and deposition of suspended matter (4). Even in digesters receiving conventional non-chemical sludges, mixing is often inadequate (1). Unfortunately, little design guidance is available to define "adequate" mixing. In general, strong mixing can be achieved if the power dissipated in the tank is 5-8 W/m³ (0.2-0.3 hp/1,000 ft³) of digester volume. Velocity gradients of 50-85 m/sec/m have been recommended (4). Use of values in the high end of these ranges is prudent when digesting chemical sludges.

Poor liquid-solids separation in secondary digesters may occur when handling alum sludges. This results in a high BOD and SS load in the return supernatant and potentially a thin sludge for dewatering or

disposal. Polymer addition to the feed sludge or in conjunction with the primary coagulant may be of some value. Poor liquid-solids separation may be the result of digester "crowding." Thickening of raw sludges can reduce the volume for digestion. More rapid removal of sludge from the digester can also reduce crowding. However, this requires sufficient dewatering and disposal capability, and consideration must be given to reduced detention time and the potential impacts on sludge stability. In some cases, changing the primary coagulant or modifying the point of chemical addition can improve liquid-solids separation (1).

5.3.5 Conditioning

5.3.5.1 Chemical Conditioning

Chemical conditioning is frequently used to improve the dewaterability of organic sludges. Addition of alum to primary or secondary processes will likely change the dewatering characteristics of the sludges and the resulting conditioning requirements. Unfortunately, it is virtually impossible to provide specific design criteria for chemical conditioning of alum sludges due to the numerous factors which affect conditioning requirements and dewatering characteristics. Polymers have become increasingly popular as conditioning agents for dewatering combinations of chemical and organic sludges, although ferric chloride and ferric chloride plus lime are also used for this purpose.

The only generalization that can be made regarding conditioning requirements of chemical-organic sludges is that conditioner dosages can be expected to be higher than for conventional non-chemical sludges. Farrell predicted that costs of conditioning may be as much as 40 percent higher when alum is used as a coagulant compared to the baseline cost of conditioning anaerobically digested primary plus waste activated sludge without chemical addition to the wastewater (25). In some cases, conditioning costs per unit weight of dry solids have more than tripled (2). In general, sludges should be conditioned and dewatered when "fresh," as storage has been shown to greatly increase conditioning requirements and adversely affect dewatering characteristics (26).

For conventional non-chemical sludges, typical conditioner dosages of ferric chloride and lime are 20-63 kg/Mg (40-125 lb/ton) and 75-277 kg/Mg (150-550 lb/ton) of dry solids, respectively. For polymer, typical dosages are 0.3-5 kg/Mg (0.5-10 lb/ton) of dry solids (4). Novak and O'Brien studied polymer conditioning of chemical sludges in the laboratory. It was found that for the near neutral pH range, anionic polymers with a range of 15-30 percent hydrolysis require the least dose and significantly reduce specific resistance. For neutral and slightly acidic sludges, cationic polymers function effectively, although dosage requirements are greater

than for anionic and nonionic polymers. For conditioning of sludges prior to vacuum or pressure filtration, results indicated that the benefit of polymer conditioning is improvement in the filtering rate and not by increasing the cake solids concentration (27).

Proper dosages of chemical conditioners can only be determined through trial and error procedures using the actual sludge to be dewatered. Where possible, results of laboratory tests should be confirmed at pilot- or full-scale.

5.3.5.2 Thermal Conditioning

Thermal conditioning is a process by which sludge is subjected to temperatures of 177-240°C (350-465°F) in a reaction vessel at pressures of 1,720-2,760 kN/m² (250-400 psig) for a period of 15-40 minutes. One variation of the process involves injection of a small amount of air into the system. Thermal conditioning changes the cellular structure of the sludge, allowing the resulting material to be readily thickened and dewatered.

There is evidence in the literature to indicate that sludges resulting from chemical precipitation of phosphorus have an adverse impact on thermal sludge conditioning processes (1,28). At Midland, Michigan, ferric chloride addition to primary treatment was initiated to reduce effluent phosphorus concentrations. When no ferric chloride was added, the thermally conditioned sludge thickened to 13 percent solids. Upon addition of 19 mg/l FeCl₃ to the raw wastewater, thickening to only 9 percent solids was possible. However, by increasing the temperature of the thermal conditioning unit from 185°C (365°F) to 202°C (395°F), the sludge could be thickened to 22.5 percent solids. Full-scale trials with alum addition were also conducted. When operated at 202°C (395°F), the thermal conditioning unit produced a sludge which could only be thickened to 16 percent solids. Percent solids of the vacuum filter cake was 41 percent for the thermally conditioned alum sludge vs. 56 percent for the thermally conditioned ferric chloride sludge.

5.3.5.3 Freezing

Subjecting alum sludge to freezing conditions is another conditioning technique that may be applicable in cold climates. In experiments at Ely, Minnesota, it was found that alum sludges with solids concentrations of 0.25-0.32, after subjecting to a natural freeze thaw cycle, dewatered to 16.8-18 percent solids. Freezing rates of alum sludges were similar to that of water (29). Reed *et al* have proposed design criteria and procedures for sludge dewatering systems utilizing natural freezing in cold climate (30).

5.3.6 Dewatering

The following sections address dewatering systems for sludges resulting from chemical phosphorus

removal processes. Much of the research on dewatering of chemical sludges was conducted during the 1970s. Since that time, there have been significant developments in sludge dewatering equipment, including displacement of traditional vacuum filters with belt filter presses, and improvements in centrifuge design. Unfortunately, little data have been published recently on the performance of these devices when processing chemical-laden sludges. For this reason, it is recommended that the engineer approach such designs cautiously. Manufacturers should be contacted for additional data and a list of facilities that are dewatering chemical sludges prior to process or equipment selection.

5.3.6.1 Drying Beds

Of the 174 plants practicing phosphorus removal and which respond to the EPA survey, 58 (33 percent) of the plants dewatered their sludges on sand drying beds. Twenty of these plants used alum for phosphorus precipitation, three of which reported problems with dewatering. Problems ranged from longer drying times to complete failure to dewater, requiring use of alternative dewatering methods (1).

Novak and Montgomery investigated sand bed dewatering on a variety of chemical sludges from water treatment plants (31).

Little design information is available for sand bed dewatering of sludges from plants using alum for phosphorus removal. For that matter, there is little published design criteria for any sludge other than anaerobically digested. For anaerobically digested, unconditioned sludge, recommended solids loadings are 134 kg/m²/yr (27.3 lb/ft²/yr) for "primary", 110 kg/m²/yr (22.4 lb/ft²/yr) for "primary plus chemicals" and "primary plus low-rate trickling filter", and 73 kg/m²/yr (14.9 lb/ft²/yr) for "primary plus waste activated sludge" (4,32). Other design criteria are in terms of bed surface area per capita, which is of limited value and has no rational design basis.

Addition of polymer to sludge prior to application to drying beds has been shown to improve dewatering and shorten drying time (1). Conclusions from the EPA survey suggested the following modifications to improve drying bed performance (1):

1. Improving performance of upstream facilities (e.g. thickeners, digesters).
2. Adding chemicals to improve dewatering characteristics.
3. Optimizing sludge loading rates and bed turnover rates.
4. Changing the drying bed filter material.

5. Covering open beds where climatic conditions adversely affect performance.

5.3.6.2 Vacuum Filtration

Many existing wastewater treatment plants employ vacuum filtration for sludge dewatering. In the EPA survey, 21 percent of the plants removing phosphorus used vacuum filtration. Most plants reported significant increases in sludge generation rates upon alum addition for phosphorus removal, which necessitated longer operating times for the vacuum filters (1).

Reports in the literature vary widely regarding the impact of alum addition to the wastewater on vacuum filter dewatering. Laboratory Buchner funnel tests showed that alum addition to secondary treatment produced a waste activated sludge that was easier to dewater, measured as specific resistance. Improved dewaterability was also found for combined raw primary and alum waste activated sludge. These sludges also required less ferric chloride to condition (22). In another laboratory study, specific resistance and filter leaf tests were conducted on aerobically digested waste activated sludge with and without alum addition prior to secondary clarification. Results showed that the alum digested sludge filtered slightly better than the control sludge. Filter leaf tests indicated that the conditioned alum waste activated sludge could be dewatered from 1.6 percent TSS to a cake concentration of 16 percent TSS at a filtration rate of 15 kg/m²/hr (3.0 lb/ft²/hr) (17).

In laboratory studies investigating dewatering of primary sludges, alum-primary sludge derived by addition of 200 mg/l alum to raw wastewater exhibited lower resistivity and capillary suction time (improved dewatering) compared to the control; however, filter leaf cake solids were 24.0 percent for the alum-primary sludge vs. 32.5 percent for the control. Required filter area for the alum sludge was projected to be 2.73 times that required for primary sludge without chemicals (33). Required filter area was calculated based on estimated sludge production rate as well as dewatering characteristics (Buchner funnel data) (32).

Pilot studies were conducted by Envirotech Corporation on thickening and dewatering of chemical primary sludges. Data from vacuum filtration of alum-primary and primary sludge indicated that alum addition to the raw wastewater adversely affected vacuum filtration of the resulting sludge. The required lime conditioning dose increased and the filtration rate decreased as the alum dosage increased. Cake solids were approximately the same for the alum dosages investigated, although were lower than primary sludge with no chemicals. Vacuum filtration performance for the alum-primary sludge is summarized in Table 5-5 (10).

Table 5-5. Summary of Vacuum Filter Performance for Alum-Primary Sludge.

Criterion	Percent P Removal with Alum	
	80	90
Lime conditioning dose, percent	25	25
Filter yield, kg/m ² /hr ¹	13.5	8.6
Cake solids content, percent ²	25.5	25.6
Solids capture, percent	97-99	97-99

¹ Excluding chemicals.

² Including chemicals.

At the 91,000-m³/d (24-mgd) West Windsor primary plant in Windsor, Ontario, effluent phosphorus levels of 1 mg/l were achieved with 90 mg/l alum and 0.4 mg/l polymer. Sludge production (dry solids) increased from 115 to 259 kg/1,000 m³ (960 to 2,160 lb/10⁶ gal), while the solids content of the primary sludge dropped from 11.5 to 7.6 percent. Vacuum filter yield dropped from 55 to 28 kg/m²/hr (11.3 to 5.8 lb/ft²/hr), and cake solids were reduced from 31.1 to 19.2 percent. Sludge conditioning with ferric chloride and lime became more difficult, increasing the costs for conditioning chemicals threefold (34).

At Windsor's 15,000-m³/d (4.0-mgd) Little River conventional activated sludge plant, 150 mg/l alum was added to the raw wastewater. Primary sludge, containing a small amount of waste activated sludge, showed a reduction in solids concentration from 6.2 to 5.7 percent, while sludge production rose from 189 to 293 kg/1,000 m³ (1,580 to 2,440 lb/10⁶ gal). Filter yield dropped from 25 to 22 kg/m²/hr (5.2 to 4.6 lb/ft²/hr). Filter cake solids were 15.9 percent when dewatering alum sludge (34).

At Lakewood, Ohio prior to alum addition to secondary treatment, average total solids in anaerobically digested sludge (vacuum filter feed) was 4.45 percent. Sludge dewatered to 23.8 percent solids. After alum addition, digested sludge feed solids concentration increased to 6.5 percent, but the dewatered cake solids dropped to 21.4 percent.

Although the dry mass of sludge generated increased from 590 Mg/yr (650 tons/yr) in 1974 to 1,650 Mg/yr (1,820 tons/yr) in 1976, vacuum filter operation and maintenance costs per unit mass of dry solids increased only 14 percent (4).

Design filtration rates and cake solids for various types of conventional sludges are shown in Table 5-6 (35). Relative vacuum filtration characteristics for various components of wastewater sludges are shown in Table 5-7. Composite characteristics are a function of the proportionate amount of each component in the total mixture. This is based on the assumption that the accumulative effects on vacuum

Table 5-6. Design Vacuum Filtration Rates for Conventional Sludges (35).

Type of Sludge	Filtration Rate kg/m ² /hr	Cake Solids percent
Raw Primary	24-48	25-30
Raw (primary + WAS)	10-24	16-24
Raw (primary + TF)	15-29	20-26
WAS	5-10	12-18
Anaerobically dig. (primary + WAS)	15-24	20-24
Anaerobically dig. (primary + TF)	20-29	20-28

Table 5-7. Component Vacuum Filtration Characteristics (31).

Sludge Component	Filtration Rate* kg/m ² /hr	Cake Solids percent
Primary	29-48	28-30
WAS	1-2	12-18
TF	7-10	12-15
AL(OH) ₃ •AlPO ₄	5-7	14-16
Fe(OH) ₃ •FePO ₄	7-10	12-15

* Values shown are indicative only of the relative effects of various components on the dewatering characteristics of a sludge mixture.

filtration are a function of the sum of the individual effects (35).

Table 5-8 lists estimated design factors for conventional and alum sludges, using vacuum filtration of anaerobically digested primary plus waste activated sludge as a baseline (2). The factors in the table can then be used to predict other yields. If no actual data are available on which to predict yields for other sludges, a baseline value of 20 kg/m²/hr (4 lb/ft²/hr) can be used (2).

5.3.6.3 Centrifugation

Several studies have investigated centrifugal dewatering of sludges from plants using aluminum salts for phosphorus control (10,17,32,36-38). In general, such sludges are amenable to centrifugation. However, decreased cake solids and deterioration in centrate quality may be expected. Polymer is effective in increasing solids capture, although often with a corresponding decrease in cake solids.

Studies by Envirotech using a pilot-scale solid bowl centrifuge showed that as the fraction of alum solids in the alum-primary sludge increased, both cake solids and maximum percent solids recovery decreased for any given hydraulic loading rate. As percent solids capture improved with polymer

Table 5-8. Design factors for Vacuum Filtration of Conventional Plus Aluminum Sludges (2).

Sludge Type	Relative Yield kg/m ² /hr	Rel. Cost of Conditioning* percent
<u>Al to primary</u>		
Digested primary + WAS	1.2	1.3
Raw primary + WAS	0.85	1.4
Raw primary + TF	1.35	1.2
Digested primary + TF	1.0	1.3
Digested primary	1.05	1.0
Raw primary	1.5	0.9
<u>Al to aeration</u>		
Raw primary + WAS	1.2	1.3
Digested primary + WAS	0.85	1.4

* Yield and cost factors related to yield and costs obtained with digested primary + WAS when no chemicals are added to wastewater.

addition, cake solids decreased. For solids captures greater than 80 percent, cake solids were typically 15 to 18 percent (10).

Baillod *et al.* investigated pilot-scale basket centrifugation of aerobically digested sludge from a contact stabilization plant with alum addition to the raw wastewater. Using the manufacturer's scale-up procedure, results indicated that a full size 1.2-m (4-ft) diameter basket centrifuge could dewater the digested alum-biological sludge at 1.6 percent TSS to a 16 percent TS cake at a rate of 159 kg dry solids/hr (350 lb/hr) with no chemical addition. Solids capture was 96 percent (17).

Canadian researchers investigated use of solid bowl and basket centrifuges for dewatering of several different sludges from pilot- and full-scale facilities. Increasing polymer dosages at the solid bowl centrifuge resulted in increased solids recovery for all sludges tested with the basket centrifuge. Recoveries in excess of 90 percent were possible without polymer addition for all waste activated sludges investigated. However, for the anaerobically digested alum sludge, polymer addition was necessary to obtain high solids recoveries. For alum waste activated sludges, an optimum polymer dosage of 1 kg/Mg (2 lb/ton) was suggested, resulting in cake solids concentrations of 10-11 percent. Waste activated sludge with alum dewatered slightly better than the control sludge. It was also found that particle size impacted centrifuge performance, with decreasing particle size resulting in decreased solids recovery and cake solids (33,36,37).

Mininni *et al.* investigated dewatering of aerobically digested waste activated sludge from a plant with no primary clarification. When alum was added to the raw wastewater, centrifuge cake solids dropped from

15.8 to 11.3 percent. Polymer was used as a conditioning agent in both cases, at dosages of 0.3-0.5 percent dry weight. Conditioner dosage increased during alum addition. The most significant machine variables affecting cake solids content were beach residence time and liquid ring height (38). It was estimated that costs for sludge conditioning, dewatering, and disposal (including amortization of capital) would increase by 63-74 percent, depending on plant size, when alum was added to the raw wastewater for phosphorus removal (39).

There are many variables which affect performance of centrifuges. Some of the more important process variables include source and type of sludge, feed solids content, percentage of chemical solids, feed rate, and conditioner dosage. Machine variables for a solid bowl centrifuge include bowl design, bowl speed, pool volume, conveyor speed, and conveyor pitch (35).

Because of the variability in sludge characteristics, pilot testing is recommended where feasible. Procedures are available from the various manufacturers for scale-up of pilot-test results.

5.3.6.4 Pressure Filtration

Less than 5 percent of the phosphorus removal plants contacted in the EPA survey practiced pressure filtration for sludge dewatering. None of these facilities used alum as the primary coagulant for phosphorus removal (1). Plate and frame filter presses are often used when it is desirable to produce a cake with a high solids content, as in preparing sludge for incineration.

Information on pressure filtration of sludges from plants using alum for phosphorus removal is scant. Envirotech evaluated pressure filtration during the investigation of chemical-primary sludge dewatering. Results for alum-primary and primary sludge are summarized below (10):

1. Filtration rates for alum-primary sludge were approximately double those for primary sludge. Filtration rates for alum-primary sludges typically are 1.0-4.5 kg/m²/hr (0.2-0.9 lb/ft²/hr), depending on conditioner (lime) dosage and cake thickness.
2. Cake solids were higher for primary sludge, 20-40 percent vs. 15-30 percent for alum-primary sludge.
3. Conditioning requirements using lime were 24-25 percent by weight for alum-primary sludge vs. 37-65 percent for primary sludges
4. Decreasing the cake thickness significantly increased filtration rate and cake solids for the primary sludge. This effect was less pronounced

as the percentage of alum solids in the sludge increased.

Minnini *et al.* studied pressure filtration of waste activated sludge (no primary clarification) using aluminum chlorohydrate as the conditioning agent. When alum was used for phosphorus removal, cake solids concentrations averaged 24.5 percent. With no chemical addition to the wastewater, cake solids concentrations averaged 32.1 percent. Conditioner dosages ranged from 1.1 to 1.5 percent by weight as Al_2O_3 . Operating pressure in the filter press averaged 6 kg/cm² (85 psi), with cycle times of approximately 3 hours (38,39).

Additional information on experiences with pressure filtration of chemical sludges may be found in Sections 5.4 and 5.5.

5.3.6.5 Belt Press Filtration

Belt filter presses have become quite popular in recent years for dewatering sewage sludges, since they are capable of producing dryer cakes than vacuum filters or centrifuges, and are less costly than plate and frame filter presses. Unfortunately, there is little information on their use at plants removing phosphorus by chemical addition.

The EPA survey makes reference to the use of a belt filter press for dewatering alum waste activated sludge at Westfield, New York (1). Since the plant was designed for phosphorus removal using alum, it is not possible to compare belt filter press performance without alum addition. The waste activated sludge is reported to be difficult to dewater, as there is no primary treatment. Cake produced by the belt filter press has an average total solids content of 11.5 percent. Polymer conditioning with polymer was found to be more successful than conditioning with ferric chloride and lime (1).

Data on performance of belt filter presses for dewatering of conventional sludges may be found in the EPA Process Design Manual on Sludge Treatment and Disposal (4).

5.3.7 Incineration

Thirteen percent of phosphorus removal plants which responded to the EPA survey incinerated their chemical sludges. Of these 22 plants, 6 reported significant impacts on incineration as a result of chemical addition for phosphorus removal. All plants incinerated a combination of primary and secondary sludges. Three of the plants that reported problems used alum for phosphorus precipitation. These were Warren, Michigan; Coloma, Michigan; and Richardson, Texas (1).

Critical variables for incineration of sewage sludge include:

1. Moisture content of the sludge,
2. Calorific value of the sludge, and
3. Relative proportion of volatile and inert material.

It is evident that the addition of alum for phosphorus removal can significantly impact these variables due to deterioration in sludge dewaterability (higher cake moisture content), and a higher concentration of inert chemical solids, which reduces the calorific value on a unit mass basis, thus increasing auxiliary fuel requirements. Table 5-9 lists the potential problems associated with chemical sludge incineration, and recommended solutions (40).

Table 5-9. Potential Problems with Chemical Sludge Incineration (40).

Problem	Solution
Greater sludge volume	Increase incinerator capacity or run time.
Lower caloric value	Increase supplementary fuel requirements.
Increased cake moisture content	Improve dewatering by: 1) modifying dewatering operation or sludge conditioning; 2) changing primary precipitating chemical or point of addition; 3) changing dewatering equipment or Increase supplemental fuel requirements.
Formation of clinkers	Decrease incineration temperature to below flash point; decrease residence time.
More inert solids	Increase ash disposal capacity.

Specific information regarding the impacts of alum addition on incineration processes is not available, although additional information on the impacts of other chemicals such as ferric chloride and lime may be found in Sections 5.4 and 5.5.

5.3.8 Disposal

Implementation of phosphorus removal by addition of alum to the wastewater has several implications regarding sludge disposal. The most significant impact is on the volumes of sludge for disposal. As discussed earlier, addition of alum for phosphorus removal will likely cause a significant increase in both the mass and volume of sludge to be transported and disposed. This will result in increased disposal costs.

Another concern is the increased metal content of sludges due to addition of precipitant. Addition of alum will increase the metal content of the sludge. Considerable research has been conducted in Canada regarding the potential impacts on crop yield,

organic matter degradation, nitrogen availability, and other soil parameters (41-44).

From review of the available literature on land application of alum sludges, it appears that addition of alum for phosphorus removal does not adversely affect the agricultural value of the resulting sludge compared to non-chemical sludges. Although some reduction of crop yield was noted with alum sludge in the Canadian studies, this was believed to be due to the high content of petroleum hydrocarbons, which was 14-30 times higher than the other sludges investigated (44). Kirkham and Dotson found that the presence of aluminum and iron phosphate precipitates did not affect the growth of barley in loam soil irrigated with wet primary sludges (45).

Increased metal content in alum sludges may impact sludge loading rates on agricultural land. Heavy metal analyses must be conducted on a case-by-case basis for the sludge to be applied. Application rates, whether governed by nitrogen or heavy metal loadings, will be determined by state and federal regulations.

Alum addition is unlikely to affect the suitability of a sludge for disposal at a sanitary landfill. However, many landfills require that organic materials meet a maximum moisture content criterion. As alum addition may increase moisture content in cakes resulting from dewatering processes, this may limit the suitability for landfill disposal.

5.4 Sludge Derived from Addition of Iron Salts

5.4.1 Sludge Characteristics

Iron salts may be employed for phosphorus removal by addition to primary, secondary, or tertiary treatment processes. Of those plants that used iron and responded to the EPA survey, 32 percent added iron to primary treatment, and 57 percent added iron to secondary treatment. A small number of plants (3 percent) added iron to tertiary processes, while several plants used iron in combination with other chemicals such as lime. Four percent of the plants added iron to both primary and secondary treatment (1). Iron salts are employed more frequently on a percentage basis than aluminum salts in primary treatment.

Knowledge of sludge characteristics, as defined by conventional characterization parameters, is of little value in predicting the amenability of iron sludges to thickening and dewatering operations, since so many variables interact to affect performance. As discussed in Section 5.3.1 design of sludge handling systems must be based on laboratory tests at a minimum, and preferably on pilot- or full-scale tests.

5.4.2 Sludge Generation Rates

As with aluminum addition, increased solids production during iron addition to wastewater results from (2):

1. Formation and removal of chemical solids such as metal phosphates and metal hydroxides,
2. Improved removals of organic solids during clarification, and
3. Removal of dissolved solids.

Although procedures are available to estimate the quantities of additional solids resulting from chemical addition, these estimates may not be accurate for a particular wastewater and treatment plant. For upgrading of existing plants, full-scale trials under controlled conditions will provide the best data regarding sludge production. For new facilities, pilot-plant tests are preferred for accurate prediction of sludge production.

The procedure for estimating sludge generation is very similar to that described in Section 5.3.2 for alum sludge, and consists of determining production of chemical solids, generation of additional solids removed during clarification, and removal of dissolved solids.

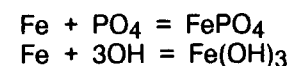
a. Chemical Sludge

This calculation assumes that the iron reacts with phosphorus compounds first, and that excess iron forms iron hydroxide.

Given:

$$\begin{aligned}P_{in} &= 8 \text{ mg/l} \\P_{out} &= 1 \text{ mg/l} \\Fe:P \text{ dose} &= 2.2:1 \\Atomic \text{ weight of Fe} &= 56 \\Atomic \text{ weight of P} &= 31 \\Atomic \text{ weight of FePO}_4 &= 151 \\Atomic \text{ weight of Fe(OH)}_3 &= 107 \\Fe \text{ dose} &= 2.2 \times 8 \text{ mg/l} \times (56/31) \\&= 32 \text{ mg/l as Fe}\end{aligned}$$

Stoichiometry:



$$(8 - 1 \text{ mg/l P}) \div 31 = 0.23 \text{ mmole added/l FePO}_4 \text{ produced}$$

$$32 \text{ mg/l Fe} \div 56 = 0.57 \text{ mmole/l Fe added}$$

$$0.57 - 0.23 = 0.34 \text{ mmole/l Fe in excess to Fe(OH)}_3$$

FePO₄ sludge: 0.23 mmole/l x 151 = 34.7 mg/l
FePO₄

Fe(OH)₃ sludge: 0.34 mmole/l x 107 = 36.4 mg/l
Fe(OH)₃

Total chemical sludge: 34.7 mg/l FePO₄
36.4 mg/l Fe(OH)₃
71.1 mg/l Fe sludge

Chemical sludge produced = 71.1 mg sludge/liter of wastewater treated (592 lb/10⁶ gallons). Adjust estimate by factor of 1.35 to account for additional chemical solids not predicted by equation (2):

Design estimate of chemical sludge production:

1.35 x 71.1 mg/l = 96.0 mg sludge/liter of wastewater (800 lb/10⁶ gallons)

b. Sludge from improved removal of suspended solids (primary clarifier)

Given:

SS_{in} = 200 mg/l

SS removal efficiency (no chemicals) = 50 percent

SS removal efficiency (with iron) = 75 percent

Additional Sludge Generated:

(0.75 - 0.50) x 200 mg/l = 50 mg sludge/liter of wastewater treated (417 lb/10⁶ gallons)

c. Sludge from removal of dissolved solids

Additional sludge mass (assuming 30% removal of soluble organics using iron salts)

= STOC_{in} (mg/l) x 0.30 x 2.5 x 1.18

= SCOD_{in} (mg/l) x 0.30 x 1.1 x 1.18

= SBOD_{in} (mg/l) x 0.30 x 1.6 x 1.18

The latter two equations are only applicable to influents prior to biological oxidation processes.

Given:

STOC_{in} = 50 mg/l

Additional sludge generated:

50 mg/l x 0.30 x 2.5 x 1.18 = 44.2 mg sludge/liter of wastewater (369 lb/10⁶ gallons)

If iron salts are added to primary treatment, improved BOD removals will result in reduced loadings to secondary biological processes, and thus reduced secondary sludge production. This must be accounted for in calculating estimates of total plant sludge production. Where iron salts are added to

secondary processes, additional sludge resulting from improved removal of SS in the secondary clarifier will probably be small for well operated activated sludge plants. However, for plants producing effluents with SS concentrations greater than 25-30 mg/l, an improved effluent quality from iron salt addition may result in significant quantities of additional sludge.

A design example is shown in Section 5.3.2 for calculating sludge production with and without alum addition. The same approach can be used for iron salt addition. The major difference is in the stoichiometry used to predict chemical sludge production.

5.4.3 Thickening

5.4.3.1 Gravity Thickening

As with alum sludges, reported impacts of primary or secondary iron addition on sludge thickening characteristics have been variable. In the survey of phosphorus removal plants that was conducted for EPA, respondents indicated both positive and negative impacts on thickener performance. Unfortunately, it was impossible to determine from the information provided in this report whether negative impacts were due directly to the addition of iron or to the additional sludge volumes which may have overloaded the thickener. Because of the many factors which affect thickening characteristics of sludges, thickening tests should be conducted on the actual sludges if available. For retrofit applications, full-scale trials should be conducted to generate representative samples of sludge upon which proper thickening tests can be run. Design criteria can then be established with confidence.

For design of gravity thickeners for new facilities, the designer does not have the benefit of having existing sludges on which to conduct thickening tests, and must rely on published design guidelines. Table 5-10 provides design criteria for thickeners receiving various types and combinations of iron and non-chemical sludges.

In addition to mass loading criteria shown in Table 5-10, hydraulic loading must also be considered in the design of gravity thickeners. For primary sludges, typical maximum overflow rates range from 1,000-1,200 l/m²/hr (25-30 gal/ft²/hr). However, for waste activated sludge or combinations of primary and waste activated sludges, hydraulic loadings rates should be considerably lower, 160-320 l/m²/hr (4-8 gal/ft²/hr) (4).

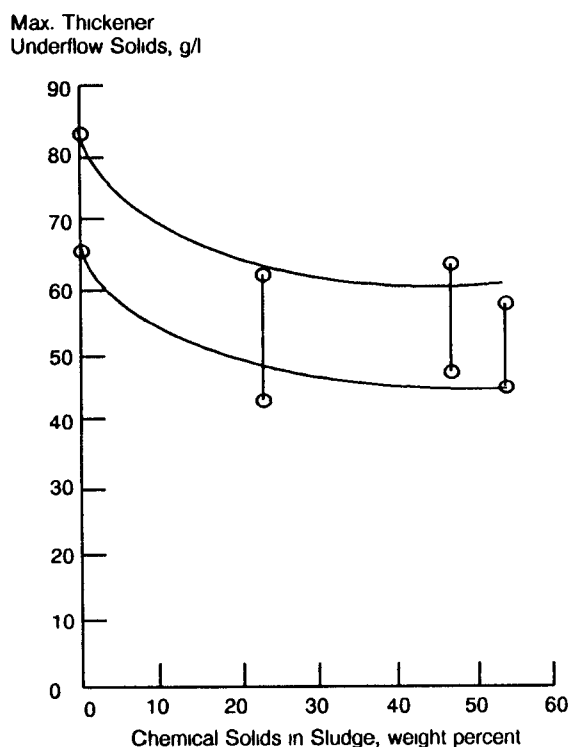
In pilot-studies of ferric-primary sludges, it was found that underflow solids concentration decreased as the amount of chemical solids in the feed sludge increased (10). This is shown graphically in Figure 5-3. Ferric-primary sludge was found to exhibit superior thickening characteristics to alum-primary

Table 5-10. Typical Gravity Thickener Design Criteria.

Sludge Type	Influent Solids Conc. percent	Expected Underflow Conc. percent	Mass Loading kg/m ² /hr	Ref.
Primary	2-7	5-10	3.9-5.9	9
Primary w/Fe*	1.8-5.2	2.2-6.4	0.3-1.3	10
Primary + WAS	0.5-1.5	4-6	1.0-2.9	8
Primary + TF	2-6	5-9	2.5-4.2	8
(Primary w/Fe) + WAS	1.8	3.6	1.3	8
(Primary w/Fe) + TF	0.4-0.6	6.5-8.5	2.9-4.2	8
Primary + (WAS w/Fe)	1.5	3	1.3	8
WAS	0.5-1.5	2-3	0.5-1.5	9
WAS w/Fe	0.5-1.5	1.5-2.0	0.6-0.8	11
Tertiary w/Fe	0.5-1.5	3-4	0.4-2.1	8

* Data reflects use of anionic polymer to assist in thickening.

Figure 5-3 Range of thickener operating periods for ferric-primary sludge (10).



sludge (see Figure 5-4). Table 5-11 shows a summary of the results at a loading rate of approximately 0.83 kg/m²/hr (0.17 lb/ft²/hr).

Canadian studies on mineral addition to an extended aeration plant found that addition of ferric chloride deteriorated the settleability of the mixed liquor

Figure 5-4 Range of thickener operating periods for alum-primary sludge (10).

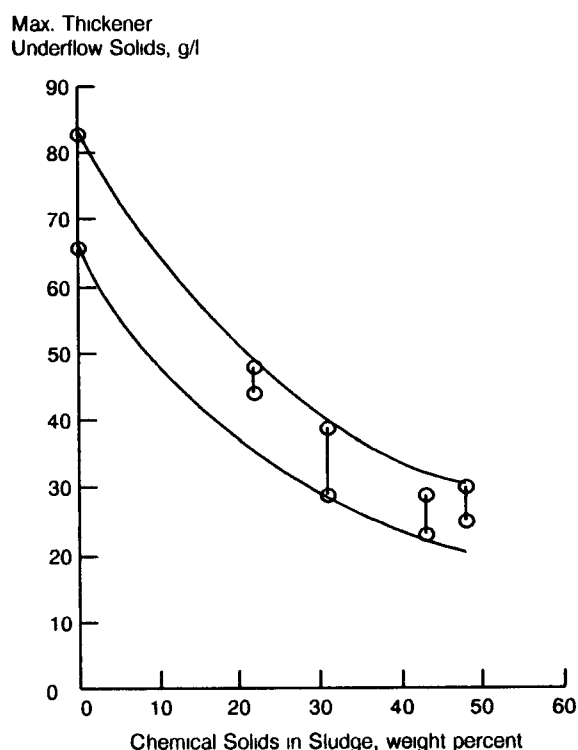


Table 5-11 Effect of Phosphorus Removal on Gravity Thickening Properties of Alum-Primary and Ferric-Primary Sludge (10).

T-P Removal percent	Alum-Primary Sludge		Ferric-Primary Sludge	
	Chemical Sludge Weight percent	Underflow TS percent	Chemical Sludge Weight percent	Underflow TS percent
80	18	4.7	22	5.5
90	23	4.1	28	5.4
95	32	3.3	3.8	5.3

Basis: Raw wastewater with 100 mg/l TSS and TP of 5 mg/l.
Solids loading rate: 0.83 kg/m²/hr (0.17 lb/ft²/hr).

suspended solids. Secondary clarifier underflow sludge was 1.1-1.3 percent solids with iron addition to the aerator vs. 1.7-1.8 percent with alum addition and 2.0 percent with no chemical addition (32).

In general, limited data are available on thickening characteristics of combined primary and secondary sludges with iron addition to the secondary process. Lower thickened sludge concentrations may be expected, and solids loading rates are generally lower than for combined sludges without chemicals. Polymer addition to the influent sludge may improve thickener performance.

5.4.3.2 Flotation Thickening

Flotation thickening is most often applied to waste activated sludges. Seldom is it used for thickening primary or combined primary-secondary sludges. In the EPA survey of plants removing phosphorus, 6 percent of the facilities employed flotation thickening. Of the four plants which flotation-thickened iron-secondary sludge, three reported no impact of iron addition on thickener performance. The fourth plant, which aerobically digested the iron-secondary sludge prior to thickening, had always treated iron sludge and could not make a comparison with non-chemical sludge (1). One plant reported use of flotation thickening for a combination of iron-primary and aerobically digested secondary sludges.

Table 5-12 summarizes the results reported from the six plants responding to the EPA survey. For iron-secondary sludges, the maximum thickened solids concentration achieved was approximately 5 percent (1).

Table 5-12. Performance of Flotation Thickeners for Treating Iron Sludges (1)

Type of Sludge	Feed Solids Conc. percent	Thickened Solids Conc. percent	Comments
Iron-secondary	0.9	5.1	Polymer added.
Iron-secondary	1.0-1.5	3.5-4.0	Polymer used when loadings high.
Iron-secondary	1.0	2.5	Polymer did not improve.
Iron-secondary (aer. digested)	1.5	2.5	Plant has no primary treatment.
Iron-primary + aer. digested secondary	-	5-6	Both cationic and anionic polymer added.

For flotation thickening of non-chemical activated sludge, typical design loadings are 2.4-3.9 kg/m²/hr (0.5-0.8 lb/ft²/hr without polymer and 4.9-7.3 kg/m²/hr (1.0-1.5 lb/ft²/hr) with polymer (8). Hydraulic loading rates are typically 29-117 m³/m²/s (0.5-2 gpm/ft²) (8). For iron-activated sludge, solids loadings of 10-24 kg/m²/hr (2-5 lb/ft²/hr) with polymer have been recommended (11). Recommended hydraulic loading rates are 58-88 m³/m²/s (1.0-1.5 gpm/ft²) (11).

5.4.4 Stabilization

5.4.4.1 Aerobic Digestion

Twenty-four percent of the phosphorus-removal plants contacted in the EPA survey reported use of aerobic digestion for chemical sludge stabilization (1). Few problems were reported, although several plants indicated problems due to increased volumes of

sludge that exceeded the design capacity of the digesters.

A Canadian laboratory study investigated the impact of chemical addition to the aeration basin on the aerobic digestion of the resulting waste activated sludge. The major conclusion was that the performance of the aerobic digestion process did not differ in any practical degree when iron or aluminum precipitates were present in the sludge (18). In addition, the release of soluble organic carbon and nutrients into the liquid phase was not enhanced by the presence of iron and aluminum precipitates. Batch digester operation resulted in greater destruction of volatile solids and a lower oxygen uptake rate than semi-continuous operation. However, semi-continuous operation at a loading rate of 1 kg volatile solids/m³/d (0.06 lb/ft³/d) provided digested sludges with better supernatant quality as well as superior settling and dewatering characteristics (18).

Relatively little information is available regarding the impact of iron salt addition to wastewater on the aerobic digestion of the resulting sludges. More information is available on alum sludges (1,17-20). Dick suggests that "Chemical precipitation of phosphorus would be expected to affect aerobic stabilization processes (aerobic digestion and composting) the same way it affects anaerobic processes. That is, changes in the amount of organic solids removed, their concentration in sludge, and their availability in sludge would influence performance (13)."

5.4.4.2 Anaerobic Digestion

Anaerobic digestion is a common sludge stabilization technique, particularly for plants larger than 19,000 m³/d (5 mgd). In the EPA survey of plants removing phosphorus, 5.6 percent of the plants employed anaerobic digestion. Of these, 22 percent reported that chemical addition was having a significant impact on their digestion process. Negative impacts reported included (1):

1. Increased energy requirements for sludge mixing, pumping, and heating,
2. Difficulty in achieving adequate digester mixing and heating,
3. Increased labor requirements for sludge pumping,
4. Poor solids - liquid separation, and
5. Reduction in digester efficiency.

Process-related problems reported with iron sludges included poor digestion, accompanied by decreased digester pH, reduced gas production, and/or decreased gas production; digester upsets characterized by loss of methane production, low pH,

and low volatile solids destruction; and digester foaming (1). On the other hand, positive effects have been reported, such as improved digester sludge settleability and supernatant quality, and increased volatile solids destruction (1). Such conflicting information is similar to that reported for alum sludge (see Section 5.3.4.2).

Malhotra *et al.* investigated anaerobic digestion of iron phosphate sludges on a laboratory scale. Conclusions from this study are as follows (46).

1. For conventional activated sludge plants using ferrous iron for phosphorus precipitation, phosphorus removal efficiency will not be drastically reduced by the return of phosphorus in anaerobic digester supernatant.
2. The pH, alkalinity of volatile acids and volatile solids destruction were similar in both the control digesters and test digesters receiving iron sludge.
3. The ferrous iron present in the feed sludge did not cause digester upset up to a maximum level of 5.5 percent Fe by weight dry solids.
4. The quantity and quality of the gas was not altered significantly with iron sludge digestion.
5. With primary sludge containing ferrous-iron precipitated phosphorus, significant uptake of total soluble phosphorus was observed during digestion. With thickened waste activated sludges containing iron-precipitated phosphorus, significant phosphorus release was observed during digestion. This was possibly due to the conversion of ferric phosphate to ferrous phosphate plus phosphate ions during anaerobic digestion.

Other laboratory studies by Dentel and Gosset concluded that chemical coagulation of organic materials with alum or ferric chloride caused a decrease in anaerobic digestibility of the resulting sludge. The effect was distinct from any effects of increased loading or differences in pH or alkalinity, and was not attributable to toxicity, increased fixed solids concentration, or phosphate limitation. Results suggested that the mechanism responsible for decreased digestibility was association of substrate with coagulant floc, rendering a portion of the organics less accessible and/or less reactive to microorganisms (23).

In another study, Gossett *et al.* found that, at ferric chloride doses of 150 mg/l to wastewater, anaerobic digester performance was 90 percent of the control; at 200 mg/l, it was 78 percent. For chemical sludges, organic nitrogen decomposition was about 50 percent less than for the control sludge, lowering ammonia production which in turn reduced alkalinity. During plant-scale studies, addition of 150 mg/l of ferric

chloride to raw wastewater resulted in a 25 percent reduction in digester gas production (24).

The most important consideration in digestion of chemical sludges is the large volumes of sludges generated during chemical precipitation. Provision of adequate digester mixing is also critical. Further discussion of these criteria is found in Section 5.3.4.2.

5.4.5 Conditioning

Chemical conditioning requirements for iron sludges can be expected to be higher than for non-chemical sludges (1). However, because of the many variables which affect conditioning requirements and dewatering characteristics, specific design criteria cannot be provided. Laboratory tests should be conducted using samples of the actual sludge to be dewatered in order to determine chemical requirements. Where possible, such results should be confirmed at pilot or full scale.

Thermal conditioning may also be affected by the presence of iron or aluminum precipitates (1,28). Because of the similarities in the results of studies on conditioning of aluminum and iron sludges, reference is made to Section 5.3.5 for further discussion of conditioning of chemical sludges.

5.4.6 Dewatering

As described in Section 5.3.6, significant advances have occurred in dewatering wastewater sludge, including introduction of the belt filter press and improvements in centrifuge design. Much of the performance data and design criteria described below resulted from work conducted in the 1970s. Little data are available on dewatering chemical-laden sludges using state-of-the-art (1987) technology. For this reason, the engineer should proceed cautiously with regard to selection and sizing of dewatering equipment. Manufacturers should be contacted regarding design criteria for dewatering chemical-laden sludges.

5.4.6.1 Drying Beds

Sand drying beds are a popular and economical technique for sludge dewatering at small wastewater treatment plants. A third of the phosphorus removal plants responding to the EPA survey used drying beds for dewatering sludge (1). Only 15 percent of the plants using drying beds reported problems with dewatering chemical sludges. Many of the problems were directly related to handling the increased volume of sludge generated by the addition of chemicals for phosphorus control. However, several plants reported that the sludge was more difficult to dewater (1).

Novak and Montgomery studied the use of sand drying beds for dewatering chemical sludges from water treatment plants (31). Conclusions from this work are discussed in Section 5.3.6.1.

Rational design criteria for other than anaerobically digested sludge are virtually non-existent (4). As discussed in Section 5.3.6.1, recommended solids loadings for anaerobically digested sludges are 73-134 kg/m²/yr (15-27 lb/ft²/yr).

Conclusions from the EPA survey suggested the following modifications to improve the performance of sand drying beds (1):

1. Improving performance of upstream facilities (e.g. thickeners, digesters).
2. Adding chemicals such as polymer to improve dewatering characteristics.
3. Optimizing sludge loading rates and bed turnover rates.
4. Changing the drying bed filter material.
5. Covering open beds where climatic conditions affect performance.

5.4.6.2 Vacuum Filtration

Twenty-one percent of the phosphorus removal plants responding to the EPA survey reported use of vacuum filters for sludge dewatering. Most plants indicated significant increases in sludge generation rates which required longer operating times for the vacuum filters (1).

As with alum sludges (Section 5.3.6.2), reports in the literature vary widely regarding the impact of iron salt addition of wastewater on vacuum filter dewatering of the resulting sludge. In the EPA survey, some plants reported significant increases in filter yield and cake solids content, while others reported decreased filter yields when dewatering iron sludges (1). Campbell and LeClair reported deteriorated dewaterability when ferric chloride was used to remove phosphorus, but improved dewaterability with alum (36). However, Mininni *et al.* found that both ferrous iron or aluminum deteriorated dewaterability, with aluminum having a more deleterious effect (38).

Envirotech Corporation conducted comprehensive pilot studies on dewatering of chemical-primary sludge (10). Vacuum filter performance, as measured by conditioner dose and filtration rate, was adversely affected by the presence of iron chemical solids in the feed sludge. Conditioner dose increased and filtration rate decreased as the proportion of iron chemical solids in the feed sludge increased. Cake solids content in dewatered ferric-primary sludge was insensitive to the proportion of chemical solids in the feed sludge; however, vacuum filter dewatering of ferric-primary sludge produced sludge cakes of higher solids than with non-chemical primary sludge. Performance of vacuum filters for dewatering iron-primary sludge was slightly better than for dewatering

alum-primary sludge. Vacuum filter performance for iron-primary sludge is summarized in Table 5-13 (10).

Table 5-13. Summary of Vacuum Filter Performance for Iron-Primary Sludge (10).

Criterion	Percent P Removal with FeCl ₃	
	80	90
Lime conditioning dose, percent	30	30
Filter yield, kg/m ² /hr ¹	10.2	8.6
Cake solids content, percent ²	34.5	34.5
Solids capture, percent	97-99	97-99

¹ Excluding chemicals.

² Including chemicals.

At Sheboygan, Wisconsin, the secondary trickling filter plant began adding ferric chloride in 1972 to the effluent of the trickling filters in order to accomplish phosphorus removal. Degritted primary sludge is blended with secondary sludge, gravity thickened, and dewatered by vacuum filtration. Comparison of vacuum filter data for 1970 and 1976 showed a decrease in filter yield from 20.3 to 12.8 kg/m²/hr (4.2 to 2.6 lb/ft²/yr).

Filter feed solids dropped from 8.6 to 7.0 percent, and cake solids were reduced from 25.5 to 21.5 percent. Filtrate quality improved from 658 ppm to 442 ppm SS. While only polymer was used for conditioning 80 percent of the time in 1970, both polymer and ferric chloride were required for conditioning after chemical phosphorus removal was implemented (1).

The City of Midland, Michigan, operates a high rate trickling filter plant with vacuum filtration of primary and secondary sludge. Prior to installation of thermal conditioning units, raw sludge was conditioned with polymers. Vacuum filter yields were 24 kg/m²/hr (5 lb/ft²/hr). When 19 mg/l FeCl₃ was added to primary treatment for phosphorus removal yields were reduced to 15 kg/m²/hr (3 lb/ft²/hr), although cake solids increased from 25.5 to 39.3 percent TS. During a period of time when thermal conditioning was on-line but when chemical phosphorus removal was not being practiced, filter yields were 39 kg/m²/hr (8 lb/ft²/hr), and cake solids were 48 percent. Upon ferric chloride addition to the primaries, filter yield dropped to 20 kg/m²/hr (4 lb/ft²/hr), and cake solids were reduced to 44 percent. Raising the operating temperature of the thermal conditioning units from 185°C (365°F) to 202°C (395°F) resulted in filter yields of 78 kg/m²/hr (16 lb/ft²/hr), and cake solids of 54 percent (1).

Design vacuum filter loading rates for various types of conventional sludges are shown in Table 5-6 (35).

Relative vacuum filtration characteristics for various components of wastewater sludges are shown in Table 5-7. Composite characteristics are a function of the proportionate amount of each component in the total mixture. This is based on the assumption that the accumulative effects on vacuum filtration are a function of the sum of the individual effects (35).

Table 5-14 lists design factors for conventional and iron sludges using vacuum filtration of anaerobically digested primary plus waste activated sludge as a baseline (2). The factors in the table can then be used to predict other yields. If no actual data are available on which to predict yields for other sludges, a baseline value of 20 kg/m²/hr (4 lb/ft²/hr) can be used (2).

Table 5-14. Design factors for Vacuum Filtration of Conventional Plus Iron Sludges (2).

Sludge Type	Relative Yield kg/m ² /hr	Rel. Cost of Conditioning* percent
<u>Fe to primary</u>		
Raw primary + WAS	1.3	1.1
Raw primary + TF	1.5	1.0
Digested primary + WAS	0.95	1.2
Digested primary + TF	1.1	1.1
Digested primary	1.2	0.8
Raw primary	1.6	0.7
<u>Fe to aeration</u>		
Raw primary + WAS	1.3	1.1
Digested primary + WAS	0.95	1.2

* Factors related to yield and costs obtained with digested primary + WAS when no chemicals are added to wastewater.

Another important design consideration is corrosion of metal components due to the corrosive nature of ferric chloride. Use of stainless steel components may be justified if ferric chloride is to be employed for phosphorus removal.

5.4.6.3 Centrifugation

Data from the literature indicate that sludges derived from processes employing iron salts for phosphorus removal are amenable to dewatering by centrifugation. However, decreased cake solids and poorer centrate quality compared to conventional non-chemical sludges may be expected in some cases.

Envirotech conducted extensive pilot studies on dewatering of chemical primary sludges. For alum-primary sludges centrifuge performance relative to polymer requirements and cake solids concentrations was adversely affected as the quantity of aluminum solids in the feed sludge increased. However, for ferric-primary sludge, cake solids concentrations increased as the amount of iron chemical solids

present in the feed sludge increased, and were significantly higher than for alum primary sludges. Polymer requirements and machine capacity to achieve a given level of solids capture were not affected as the quantity of iron chemical solids in the feed sludge increased. At total solids capture of 95 percent, centrifugal dewatering of ferric-primary sludge produced cakes of 22-25 percent solids vs. 20-21 percent solids for the control primary sludge with no iron (10).

Campbell and LeClair reported on pilot scale centrifugal dewatering of waste sludge generated by an extended aeration pilot plant (36). Using a basket centrifuge, it was found that the ferric chloride sludge was more difficult to dewater than either the alum sludge or the control sludge with no chemical. Cake solids for the iron sludge were 8.4-12.0 percent TS. With a solid bowl centrifuge, cake solids achieved with the iron sludge were essentially the same as the control sludge, but consistently lower than the alum sludge. Cake solids with iron sludge were 3.4-8.4 percent TS. Polymer addition was required to obtain solids recoveries in excess of 9.5 percent (36).

Centrifugation of anaerobically digested primary and iron-secondary from the North Toronto plant was also investigated. A feed sludge of 6.6 percent TS was dewatered to 17.3-24.0 percent TS with polymer dosages of 4.4-6.5 kg/Mg (8.8-13.0 lb/ton). Solids recoveries were 98-99 percent (33).

Mininni *et al.* investigated centrifuge dewatering of aerobically digested waste activated sludge from a plant with no primary clarification. When ferrous sulfate was added to the raw wastewater, sludge cake concentrations dropped from 15.8 percent to 13.0 percent. Cationic polymer was used as a conditioning agent at similar dosages in both cases. The most significant machine variables were beach residence time and liquid ring height (38). Estimated sludge handling costs were slightly lower for iron addition than for aluminum addition (39).

As discussed in Section 5.3.6.3, process variables affecting centrifuge performance include source and type of sludge, feed solids content, percentage of chemical solids, feed rate, and conditioner dosage. Machine variables for a solid bowl centrifuge include bowl design and speed, pool volume, conveyor speed, and conveyor pitch (35).

Pilot testing is strongly recommended where possible, as the variability in sludge characteristics makes performance impossible to predict. Procedures are available from the various manufacturers for scale-up of pilot-test results.

5.4.6.4 Pressure Filtration

Limited information is available on pressure filtration of iron sludges. In the EPA survey, plants reporting

use of plate and frame filter presses on iron sludges or combinations of iron and conventional sludges included Saline, Michigan; Kenosha, Wisconsin; and Brookfield, Wisconsin (1). At Saline, anaerobically digested iron-primary plus secondary sludge at 9 percent TS was dewatered to 41 percent TS; at Kenosha, anaerobically digested primary plus iron-secondary sludge at 4.8 percent TS was dewatered to 40 percent TS; and at Brookfield aerobically digested primary plus iron-secondary sludge at 7 to 8 percent TS was dewatered to 43 percent TS (1).

At Brookfield, Wisconsin, filter press performance improved substantially upon addition of pickle liquor (ferrous sulfate) to the aeration basins for phosphorus removal. The sludge mass feed rate (not including admixtures) increased 70 percent from 227 to 384 kg TS/hr (500 to 845 lb/hr); average run length decreased by 40 percent from 2.8 to 1.7 hr/run; and the percentage of solids fed to the filter which were recovered in the cake increased from 75 to 90 percent. The cake solids content remained essentially constant at 43 percent TS (1).

Mininni *et al.* investigated pressure filtration of waste activated sludge from a plant with no primary clarification. Upon addition of ferrous chloride to raw wastewater for phosphorus control, cake solids decreased from 32.1 to 30.0 percent TS. Aluminum chlorohydrate was used as a conditioner at dosages of 12.7 and 12.1 g $\text{Al}_2\text{O}_3/\text{kg}$ (25.4 and 24.2 lb/ton) of dry solids during the periods of no chemical addition and during ferrous sulfate addition, respectively (38,39).

For further general information on pressure filtration of sludge the reader is referred to the EPA Process Design Manual on Sludge Treatment and Disposal (4).

5.4.6.5 Belt Pressure Filtration

Many plants are now using belt filter presses for sludge dewatering, since, in general, dryer sludge cakes are possible compared to vacuum filters and centrifuges, and the devices are economically competitive with other dewatering equipment. Unfortunately, virtually no information is available in the literature regarding their performance on sludges derived from the addition of iron salts for phosphorus removal.

Performance data and design criteria for belt filter presses used for dewatering of non-chemical sludges may be found in Reference 4. Pilot-testing is recommended if belt filter presses are under consideration for dewatering of sludges from chemical phosphorus removal systems.

5.4.7 Incineration

Of 22 plants contacted in the EPA survey of phosphorus removal facilities which employed

incineration, six reported significant impacts on incineration as a result of chemical addition for phosphorus removal. Two of these plants used iron salts for phosphorus removal. These were Wyandotte, Michigan and Sheboygan, Wisconsin.

Wyandotte reported major problems with clinker buildup in the multiple hearth incinerator. This increased the time required for cleaning drop holes in the hearth, and increased wear on the rabble arms.

Sheboygan operates a fluidized bed incinerator, which receives sludge cake from a vacuum filter. Ferric chloride is added to the effluent of the trickling filters for phosphorus removal. Phosphorus removal has adversely affected incinerator capacity due to the increased moisture content of the sludge cake from the vacuum filter. Cake solids dropped from 25.5 to 21.5 percent after phosphorus removal was initiated. The volatile fraction was reduced from 73-74 percent to 65 percent. The overall impact was a reduction in incinerator feed rate from 755 to 475 kg dry solids/hr (1,660 to 1,050 lb/hr), and an increase in fuel consumption from 246 to 517 l/Mg (59 to 124 gal/ton) (1).

Other problems associated with incineration of chemical sludge have been reported at Sheboygan. Slag formation has caused plugging of tuyeres and clogging of exhaust lines, increasing the frequency of inspections and maintenance. In addition, the plant manager believes that the ferric chloride in the sludge is responsible for a high rate of corrosion of metal ductwork (1).

It is apparent from the Sheboygan experience that ferric chloride addition for phosphorus removal can impact critical incineration parameters such as: 1) cake moisture content, 2) calorific value of the sludge, and 3) relative proportion of volatile and inert material. Table 5-9 lists the potential problems associated with chemical sludge incineration, and recommended solutions (40).

5.4.8 Disposal

As discussed in Section 5.3.8 for disposal of alum sludges, the most significant impact of chemical addition on sludge disposal considerations is the increased sludge mass and volume for disposal, resulting in increased disposal costs.

Information in the literature indicates that the presence of iron precipitates does not adversely affect the agricultural value of the sludge with respect to crop yield, organic matter degradation, nitrogen availability, and other soil parameters (41-45). However, an increase in metal content of the sludge may be expected as a result of precipitation with iron. This may impact allowable loading rates to agricultural land. Heavy metal analyses must be conducted on a case-by-case basis for the sludge to be applied.

Application rates, whether governed by nitrogen or heavy metal loadings, will be determined by state regulations.

Iron addition is unlikely to affect the suitability of a sludge for disposal at a sanitary landfill. However, mass landfills require that organic materials comply with a maximum moisture content criterion. As iron addition may increase moisture content in sludge cakes from dewatering processes, this may limit the suitability for landfill disposal.

5.5 Sludges Derived from Biological Phosphorus Removal Processes

5.5.1 Characteristics

Sludges derived from biological phosphorus removal systems exhibit properties similar to conventional biological sludges. The only possible exception is sludge derived from the Phostrip process in which a portion of the total sludge results from lime addition to the anaerobic stripper vessel. Even with the Phostrip process, the volume of lime sludge is relatively small compared to the combined volume of primary and waste activated sludge.

Because of the mechanism of excess phosphorus uptake in biological phosphorus removal systems, resulting waste activated sludges tend to have higher phosphorus concentrations than conventional sludges. Typical phosphorus concentrations in waste activated sludge from the Bardenpho and A/O processes are 4-6 percent by weight vs. 2-3 percent for conventional waste activated sludges.

Resolubilization of phosphorus during anaerobic storage of lime-precipitated sludge in the Phostrip process is unlikely to occur, since the phosphorus is bound to the calcium ion. However, with A/O, Bardenpho, and other "pure" biological phosphorus removal systems, it is recommended that waste activated sludge be kept aerobic in order to prevent phosphorus solubilization.

5.5.2 Sludge Generation Rates

Sludge generation rates from biological phosphorus removal systems are not expected to be significantly different than for conventional activated sludge systems, and solids production will vary with wastewater characteristics and operational parameters such as SRT.

The Phostrip process will likely generate somewhat greater masses of sludge because of lime addition to the anaerobic stripper supernatant. Knowing the characteristics of the supernatant and the necessary pH to remove the desired quantity of phosphorus, the quantity of lime sludge produced can be estimated using the procedure described in Section 5.6.

Theoretically, some increase in sludge production would be expected for biological phosphorus removal systems due to the increased mass of phosphorus taken up by the organisms. This will be dependent on the phosphorus content of the waste activated sludge (WAS). As shown in Section 3.5.2.1, the theoretical waste activated sludge yield would increase by 8.5 percent if the phosphorus content of the WAS increased from 2 to 4 percent by weight. If the phosphorus content increased to 5 percent by weight, the theoretical mass of WAS production would increase by 13 percent. It should be noted that on a volumetric basis, the increased sludge mass may be counteracted by an improvement in settling characteristics, as SVI values of less than 80 ml/g have been reported for the Modified Bardenpho and A/O processes.

5.5.3 Thickening

Because of the potential release of phosphorus during gravity thickening of waste activated sludge from biological phosphorus removal systems, use of dissolved air flotation thickening is recommended. This would apply to the purely biological systems, and would not be a concern with lime sludges derived from treatment of the anaerobic stripper supernatant in the Phostrip process. Lime sludges can be combined with other sludges and handled by conventional sludge handling processes without special consideration for phosphorus release. With "pure" biological systems, however, choice of sludge handling processes must account for potential phosphorus resolubilization if thickened, stored, or stabilized in the absence of oxygen.

5.5.4 Stabilization

Little data are available regarding the fate of phosphorus during aerobic or anaerobic stabilization of sludges from biological phosphorus removal systems. Phosphorus resolubilization would be anticipated during anaerobic digestion. However, at Pontiac, Michigan, significant levels of phosphorus in anaerobic digester supernatant were not observed, possibly due to formation of an ammonium-magnesium-phosphate precipitate in the digester.

Phosphorus release may also be possible during aerobic digestion due to destruction and lysing of biological solids.

Because of the lack of information on this subject, consideration may have to be given to chemical treatment of digester supernatants for phosphorus removal in order to minimize return of phosphorus to the head of the plant. Further studies are needed to assess the magnitude of phosphorus release during stabilization of biological phosphorus removal sludges.

In some biological phosphorus removal systems employing long solids retention times, phosphorus-laden sludges are subjected to dewatering without

separate stabilization. The acceptability of this practice is dependent on regulations as to whether such sludges are considered stabilized or whether separate stabilization is required prior to land disposal.

5.5.5 Conditioning

Sludges from biological phosphorus removal systems are expected to have similar conditioning requirements to conventional non-phosphorus sludges. Blending of lime sludge with other sludges in the Phostrip process may reduce overall conditioning requirements.

5.5.6 Dewatering

Sludges from biological phosphorus removal systems are expected to have dewatering characteristics similar to those from conventional activated sludge systems. The lime sludge from the Phostrip process is not be expected to adversely affect dewatering, and, based on experience with dewatering of lime sludges alone, may improve dewatering characteristics when blended with other sludges. No specific design information regarding loading rates to dewatering equipment is available for biological phosphorus removal sludges. Design criteria for conventional primary and waste activated sludges should be used to size dewatering equipment for sludges from biological phosphorus removal processes if pilot- or full-scale performance data are not available.

5.5.7 Incineration

No unique problems are associated with incineration of sludges from biological phosphorus removal systems. Sludges from "pure" biological phosphorus removal processes will have volatile solids contents and Btu values similar to those of conventional biological wastewater treatment sludges. Phostrip sludges, if lime sludge from treatment of the stripper supernatant is included, may have slightly lower volatile solids contents and Btu values due to the addition of inert solids from the lime addition step. However, the overall impact is expected to be small.

5.5.8 Disposal

Sludges from biological phosphorus removal systems can be disposed of in the same manner as sludges from conventional biological systems. Higher phosphorus contents may make sludges from biological phosphorus removal systems particularly attractive for agricultural utilization.

5.6 Sludges Derived from Addition of Lime

Lime may be employed for phosphorus removal by addition to the primary clarifier or tertiary treatment process. Lime is also used to remove phosphorus from the effluent of the phosphorus stripper tanks in

the Phostrip biological phosphorus removal systems (47). The characteristics of the sludge produced are dependent on where the lime is added, whether the low lime or high lime process is used, the alkalinity of the water, and whether the various wastewater plant sludges are combined with the lime sludge before processing and in what proportions they are combined.

As discussed in Section 4.1, very few plants in the United States use lime for phosphorus removal. Many plants originally designed to use lime have abandoned the lime systems in favor of aluminum or iron salts. The major disadvantages of lime compared to metal salts are significant increase in sludge mass, and greater operation and maintenance requirements for cleaning and maintaining lime handling equipment. The following is a brief discussion of lime sludges with references provided if further detail is desired.

Two general statements about lime sludges can be made. One is that use of lime for phosphorus removal results in larger volumes of sludge before thickening than does the use of metal salts for phosphorus removal (1,48). Secondly, lime sludges generally improve the thickening and dewatering properties of wastewater treatment sludges when lime sludge and non-lime wastewater sludges are mixed (13).

Due to the fact that the nature of the sludge produced will vary with the specific wastewater being treated and the composition of the sludge, the design of the sludge handling system must be based at a minimum on laboratory tests such as settleometer tests, specific resistance, filter leaf, capillary suction time and other tests. Where possible, pilot- or full-scale sludge thickening and dewatering devices should be used to establish design criteria for full-scale sludge thickening and dewatering devices.

Specific differences expected to be found in the different lime treatment systems are listed below.

A. Low Lime Treatment - Primary Addition.

1. The sludge contains a smaller percentage of organics than does high lime treatment (49).
2. The sludge contains no magnesium hydroxide, which is a gelatinous precipitate that is difficult to settle (13,49).

B. High Lime Treatment - Primary Addition

1. Sludge removed from high lime treatment has a higher solids content than that from low lime treatment (50).
2. The sludge contains magnesium hydroxide, a gelatinous precipitate.

C. Tertiary Lime Process (11).

1. Tertiary lime sludges are similar to primary lime sludges but do not contain nearly the quantity of organic materials as primary sludge.
2. The high lime tertiary process will produce a sludge that behaves much like sludge from a water softening process. The thickening and dewaterability of this sludge decreases with increasing magnesium concentration in the sludge.
3. High alkalinity waters, (greater than 200 mg/l CaCO_3) require special consideration for thickening of the sludge as the thickening operations can become unmanageable.
4. Recycle of sludge from the settling unit improves thickening of the sludge.

Specific data on the thickening, dewatering and disposal of lime sludges are contained in the following sections. Recommended reading for primary lime sludges are articles by Minton and Carlson (48), who did a number of studies on these sludges in the early 1970's, and Parker (50,51).

Theoretical calculations of sludge generation rates appear to give good correlation with observed sludge volumes in plant studies for tertiary lime addition but underestimate the quantity for primary lime addition (48). The quantity of sludge produced by lime addition depends not only on the degree of phosphate removal desired, but also on the magnesium concentration, alkalinity and other characteristics of the wastewater (52). In fact, both the amount of sludge produced and the degree of phosphorus removal achieved by raising the pH to a given level depends on the nature of the wastewater.

Increases in the mass of sludge solids produced by lime addition systems have been reported to be double the mass produced by conventional primary and secondary treatment systems (53). Minton and Carlson have reported the mass of lime sludges to be as much as two to three times the mass of conventional primary and secondary treatment systems when lime is added to raw wastewater.

An important factor that must be kept in mind is that while the sludge mass increases, lime sludges generally thicken and dewater to high solids content and the volume of dewatered sludge that needs to be processed is not necessarily two to three times the volume of sludge from a wastewater plant without chemical phosphorus removal.

Procedures have been developed for calculating the sludge quantity produced from lime sludges. These procedures were developed for tertiary lime

applications but can be used for raw sewage or other applications. The procedure is described in detail in the 1976 EPA Phosphorus Design Manual (4).

5.7 Case Histories

5.7.1 Baltimore, Maryland (54)

The Back River Wastewater Treatment Plant provides secondary treatment of municipal-industrial wastewater from Baltimore City and Baltimore County, Maryland. Liquid stream processes include preliminary treatment, primary clarification, biological treatment by parallel trickling filters and activated sludge process, secondary clarification, and chlorination. Primary and waste activated sludges are thickened by gravity, anaerobically digested, chemically conditioned with polymer, and dewatered using vacuum filters. Phosphorus is removed by addition of iron as waste pickle liquor to the activated sludge aeration basins at a dosage of approximately 5 mg/l as Fe. Full-time addition of waste pickle liquor began in June, 1981.

Data on sludge handling operations were compared for periods with and without waste pickle liquor addition. Prior to waste pickle liquor addition, average daily sludge production was 62 Mg (69 tons)/d dry solids, and 353 Mg (390 tons)/d wet. After pickle liquor addition, average daily sludge production was 80 Mg (88 tons)/d dry solids, and 414 Mg (456 tons)/d wet. This amounted to an increase of 29 percent on a dry solids basis, and 17 percent on a wet solids basis following initiation of phosphorus removal. The smaller percentage increase in wet sludge production was due to generation of a drier vacuum filter cake as a result of waste pickle liquor addition. Cake solids increased from 17.6 percent to 19.1 percent, an improvement of 9 percent.

Coupled with an increase in vacuum filter cake solids was reduction in polymer required for sludge conditioning. Prior to iron addition, polymer dose based on dry solids, was 34 kg/Mg (68 lb/ton); after iron addition, polymer dose was 30 kg/Mg (60 lb/ton), a decrease of approximately 11 percent.

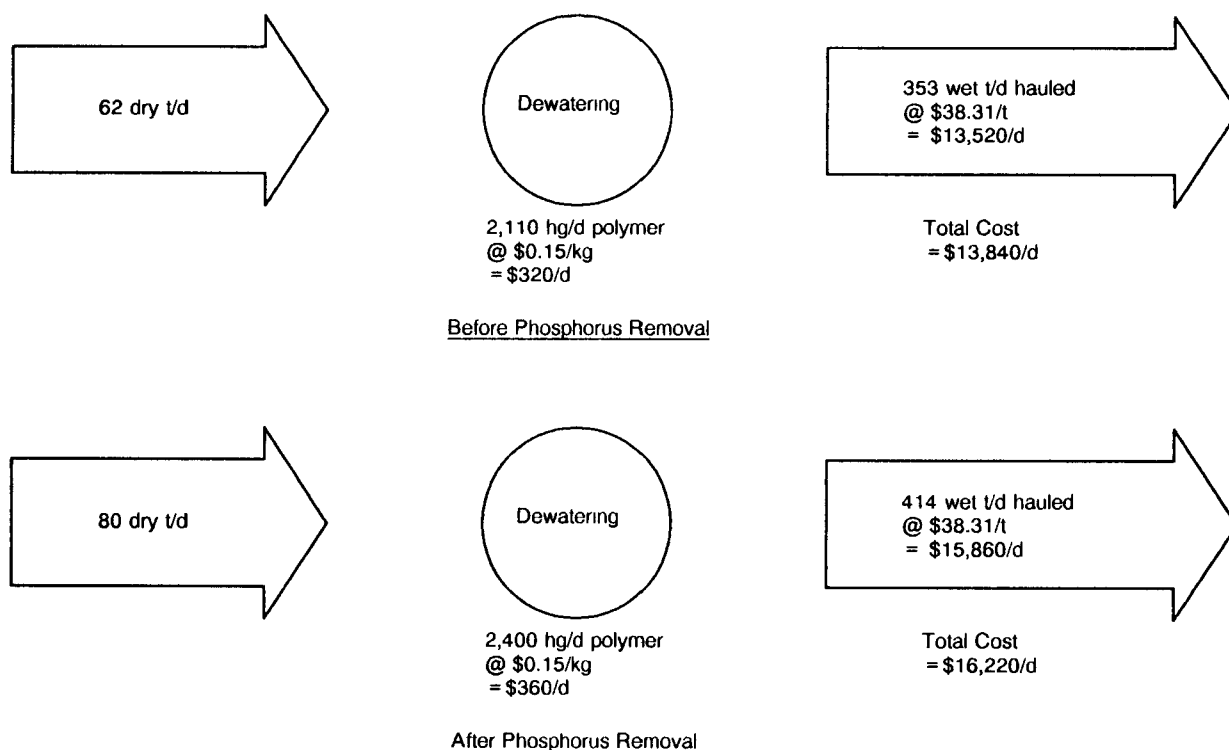
Thickening characteristics of combined raw primary and secondary sludges did not appear to change significantly as a result of pickle liquor addition. Changes were attributed to operational modifications.

Figure 5-5 summarizes the costs of sludge conditioning and hauling before and after phosphorus removal. It should be noted that these costs are only for chemical conditioning and dewatered sludge hauling, and do not account for additional operation and maintenance labor or energy requirements.

5.7.2 Lorton, Virginia (55)

The Lower Potomac Water Pollution Control Plant is a 136,000-m³/d (36-mgd) activated sludge facility

Figure 5-5 Comparison of sludge quantities, conditioning costs, and hauling costs before and after phosphorus removal; Baltimore, MD.



with advanced wastewater treatment for phosphorus removal. Liquid stream processes consist of bar screens, primary clarifier, activated sludge basins, secondary clarifier, flow equalization basins, tertiary chemical clarifiers for phosphorus precipitation, and tertiary filters. Primary sludge is dewatered and gravity thickened. Waste activated sludge is thickened by dissolved air flotation. Primary and waste activated sludges are blended in a storage tank, chemically conditioned with lime and ferric chloride, and dewatered by vacuum filtration. Dewatered sludge cake, at 16-18 percent solids, is incinerated in multiple-hearth furnaces. The tertiary chemical sludge is conditioned with lime, gravity thickened to 2-3 percent solids, conditioned with anionic polymer, and dewatered in solid bowl centrifuges. Dewatered sludge cake, at a solids content of 15-16 percent, is mixed with incinerated sludge ash and disposed of in a sanitary landfill.

Originally the plant was designed for phosphorus removal in a tertiary process using a two-stage lime and recarbonation system. Because of numerous design, operation, and maintenance problems and high operating costs, the plant was modified to remove phosphorus using ferric chloride. Ferric chloride is normally added to the plant influent, activated sludge process, and influent to tertiary clarification. Ferrous sulfate addition to the return

activated sludge was investigated during the period April through July of 1983. During this time, ferric chloride addition to the plant influent and influent to tertiary clarification was maintained. From June through October, 1984, ferrous sulfate was added to the plant influent while ferric chloride addition to the activated sludge process and tertiary clarifiers was maintained. Results of these investigations with respect to impact on solids handling are discussed below.

Ferrous sulfate addition to return activated sludge

Total suspended solids removal in the secondary clarifier improved significantly during ferrous sulfate addition to the return activated sludge. TSS removal efficiencies were 87 percent during ferrous sulfate addition and 80 percent during ferric chloride addition. Performance of dissolved air flotation thickening of waste activated sludge was slightly improved during ferrous sulfate addition, with thickened solids concentrations increasing from 4.4 to 4.9 percent by weight. However, the improved performance was not solely attributed to ferrous sulfate addition.

Ferrous sulfate addition to plant influent

Gravity thickening of primary sludge improved with addition of ferrous sulfate to the plant influent, with average thickened solids concentrations increasing from 6.6 to 8.7 percent.

Performance of vacuum filters improved during the period of ferrous sulfate addition to the plant influent. Average total solids concentration in the vacuum filter sludge cake increased from 16.4 percent during ferric chloride addition to 18.7 percent during ferrous sulfate addition. There was no change in chemical requirements for conditioning.

Another observation was that the chemical "ferric" sludge from the tertiary clarifier dewatered to a higher cake solids content when ferrous sulfate was added to the influent, increasing from 15-16 percent to approximately 19 percent. Although no explanation was provided as to the reason for the improvement, the same phenomenon was observed when ferrous sulfate addition was temporarily suspended and then restarted.

The major conclusion from the study was that, although ferrous sulfate was more costly than ferric chloride per unit of phosphorus removed, the savings resulting from improved thickening and dewatering characteristics may make its use cost-effective for this application.

5.8 Costs

No estimates have been provided for the costs incurred in handling additional sludge associated with phosphorus removal processes. This is partly due to the wide variability in sludge generation rates as a result of variation in process types, operating strategies, chemical used for phosphorus removal (if any), effluent discharge limitations, wastewater characteristics, and other factors. Costs are dependent not only on the additional quantity of sludge to be processed, but also on changes in thickening and dewatering characteristics of the sludge, types of sludge handling processes employed, existing sludge handling capacity, labor and energy costs, and available sludge disposal options.

Estimates of the additional costs for sludge handling upon implementation of phosphorus removal can only be made on a site-specific basis. Sludge generation rates can be estimated using the procedures outlined in this chapter. Laboratory tests can be used to estimate the impact of phosphorus removal on thickening and dewatering characteristics. A capacity analysis of sludge handling equipment must be conducted in order to determine if more tankage or equipment will be required to process the additional sludge. A similar analysis of manpower schedules and requirements must also be made. Finally, the overall impact on the volume of sludge to be disposed of must be assessed in order to estimate additional disposal costs. For chemical precipitation of phosphorus, cost for sludge handling and disposal is likely to account for a major portion of the additional

costs associated with phosphorus removal, and cannot be ignored.

5.9 References

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

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Terms

BOD	biochemical oxygen demand
BOD ₅	5-day BOD
COD	chemical oxygen demand
DO	dissolved oxygen
F/M	food-to-microorganism loading
gpd	gallons per day
gpm	gallons per minute
HDT	hydraulic detention time
HRT	hydraulic retention time
mgd	million gallons per day
mg/l	milligrams per liter
MLSS	mixed liquor suspended solids
MLVSS	mixed liquor volatile suspended solids
NH ₄ -N	ammonium nitrogen
NO ₃ -N	nitrate nitrogen
SBOD	soluble 5-day BOD
SDT	stripper detention time
SP	soluble phosphorus
SRT	solids retention time
SS	suspended solids
TBOD	total 5-day BOD
TKN	total Kjeldahl nitrogen
TN	TKN plus oxidized nitrogen
TP	total phosphorus
TSS	total suspended solids
VSS	volatile suspended solids