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Technical Report

Methods for Biological Phosphorus Removal:  
The Potential for Application in Massachusetts

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S. Russell Silva, Commissioner  
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## ABSTRACT

The control of eutrophication in the lakes and rivers of the Commonwealth of Massachusetts has led to the imposition of phosphorus limitations on the discharges of many publically owned treatment works, which has in turn led to an interest in alternative methods to meet these permit limits. The potential of biological phosphorus removal systems to meet these needs was investigated. Factors which influence the ability of a plant to be retrofitted to biological phosphorus removal systems were identified. Three plants were selected for use as case studies to determine which systems might be most applicable and what costs would be involved. In one plant biological phosphorus removal was not deemed practical while both of the remaining plants were found to be compatible with biological systems. In one case biological phosphorus removal using the A/O from, Air Products and Chemicals, process presented a least cost alternative, while conventional chemical precipitation was the least cost alternative at the other plant.

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## CHAPTER I

## INTRODUCTION

1.1 General In many areas of the world, significant changes in the aquatic environment have been caused by the release of excess nutrients into waterbodies. When the supply of nutrients in a given waterbody increases to a point where the ecosystem is "well nourished", the waterbody is said to be "eutrophic". Though eutrophication is a naturally occurring process which is often considered analogous to aging, man's activities have accelerated the process to the detriment of many waterbodies. Wastewater discharges, agricultural runoff and urban runoff have all been cited as sources of the added nutrients.

Accelerated eutrophication causes changes in aquatic environments that often interfere with the use of the water, detract from its natural beauty, and may reduce property values. Commonly, excess algal and larger plant growth chokes open water rendering the water nonpotable and greatly increasing filtering and other treatment costs necessary to make use of the water. As the excess vegetation decomposes, foul gases may be given off and the dissolved oxygen critical to fish and other aquatic animals is consumed (Rohlich, 1969).

Phosphorus and nitrogen have long been recognized as limiting nutrients in the eutrophication of most lakes and estuaries. In order to control eutrophication, phosphorus and nitrogen effluent



limits have been placed on wastewater discharges in many parts of the world. Examples in the United States include the Great Lakes, Lake Tahoe, the Chesapeake Bay and Tampa Bay (Weston, 1984 and Switzenbaum et al., 1981). Although more and less stringent requirements are common, typical effluent standards have been 1 mg/l phosphorus and 1 to 3 mg/l total nitrogen (Weston, 1984). Phosphorus is more commonly -- and often more stringently -- controlled than nitrogen principally because some phytoplankton species are capable of fixing atmospheric nitrogen (Hogan, 1984).

While municipal wastewater is highly variable, influent phosphorus concentrations are often in the range of 7 to 10 mg/l (Schmidtke, 1980). This represents an increase from typical values in the past as in the 1940's phosphorus concentrations were typically in the 2 to 4 mg/l range. As the current phosphorus concentration levels are well above what is required by activated sludge organisms for normal metabolism, the excess phosphorus is released from most wastewater treatment plants in the effluent unless a specific phosphorus removal system is employed.

1.2 Scope and Purpose. Chemical precipitation has traditionally been used to remove excess phosphorus in wastewater treatment. Aluminium compounds, lime, and iron salts are the most widely used precipitants for this purpose. Other processes which have demonstrated the ability to remove phosphorus include ion exchange, reverse osmosis, and other demineralization techniques (EPA, 1976).

These processes are relatively expensive and are more commonly used for wastewater renovation and reuse than for pollution control and, as such, will not be considered further. A third class of phosphorus removal processes are distinguished by their use of activated sludge microorganisms to "uptake" phosphorus in excess of stoichiometric amounts and incorporate the phosphorus into cell biomass.

Inherent disadvantages of chemical precipitation include chemical costs, the need for chemical handling and storage, difficulties in matching chemical dosage to changing flows and phosphorus concentrations, and increased sludge disposal and handling costs. Biological phosphorus removal is a relatively new technology and operational difficulties are still being evaluated. Critical factors appear to include dissolved oxygen control (particularly when nitrogen removal is required) and maintaining oxygen in the secondary clarifier sludge blanket. Both types of phosphorus removal and their advantages and disadvantages will be discussed further in chapter three.

This research has been undertaken in order to assess the potential for the use of biological phosphorus removal systems in municipal wastewater treatment plants within the Commonwealth of Massachusetts.

A number of biological phosphorus systems have been developed. Among these are the Bardenpho, A/O, Phostrip, Phoredox, UTC, and Biodenipho processes (Arvin, 1985). The first three will be

considered here as they have been the most widely studied and are presently available in the United States as follows:

Bardenpho: Eimco Process Equipment Company, P.O. Box 300, Salt Lake City, Utah.

A/O: Air Products and Chemicals Inc., P.O. Box 538, Allentown, Pennsylvania.

Phostrip: Biospherics Inc., 4928 Wyaconda Rd., Rockville, Maryland.

The major questions addressed in this study were:

1. What are the important engineering and design considerations necessary to determine the suitability of biological processes to meet Massachusetts effluent requirements?

2. Which of the proprietary processes will be the most practical for various treatment plants having different design loadings, influent characteristics, and effluent requirements?

3. How do the costs associated with biological phosphorus removal compare with the costs for conventional precipitation?

The first question attempts to define the parameters by which initial decisions concerning the suitability of a given plant for

retrofit to biological phosphorus removal can be made. Though some limitations have been suggested in the literature for each process, (Weston, 1984 and Barnard, 1983) guidelines indicating when biological methods are applicable have not been widely available. The desired result was a decision-making tool which will enable the engineer to determine when biological phosphorus removal is a reasonable alternative to investigate further and when it is not.

The second question is, in some respects, a refinement upon the first. The point was to differentiate between the three proprietary processes and determine the conditions which may favor one process over another. Each of the processes has shown different phosphorus removal rates, sludge production rates, and reactor volume requirements for a given influent load. These and other process differences may be utilized to advantage in selecting a given process for a given plant.

The third question is of obvious importance. If biological phosphorus removal is to be used in the Commonwealth, it must be economically attractive or at least competitive with conventional precipitation. Therefore, in order to assess the potential for use of these processes, a determination of relative cost must be made.

## CHAPTER II

## BACKGROUND

2.1 Theory of Biological Phosphorus Removal All living cells require phosphorus for cell synthesis, metabolism, and growth. It has been found that typical cells are 1 to 3% phosphorus by dry weight (Grady and Lim, 1980). Activated sludge systems, therefore, naturally remove some phosphorus through sludge wasting. Morgan and Fruh have found that in aerobic systems where phosphate precipitation is not occurring, the phosphorus content of the sludge is primarily a function of the ratio of phosphorus and organic matter (Arvin, 1985). Actual phosphorus removal may also be influenced by the hydraulic residence time, sludge age, sludge wasting, and the speciation of influent phosphorus (orthophosphate, polyphosphorus, and organic phosphorus) (Hogan, 1984).

The phosphorus content of the volatile suspended solids in waste activated sludge has been shown to increase from 0.7 to 2.5% phosphorus (on a dry weight basis) when the COD to P ratio decreased from 2000:1 to less than 60:1 (Arvin, 1985). This maximum phosphorus concentration in the sludge is important because sewage commonly exhibits COD:P ratios in the range of 20:1 to 100:1. Also, it has been demonstrated that phosphorus is not growth limiting to activated sludge organisms when the phosphorus content of the sludge is 1.0% or more on a dry weight basis. Therefore, typical secondary sludge having a dry weight phosphorus content of 1.5 to 2.0%, and representing 10 to 30% of the influent phosphorus, is already

exhibiting removal above stoichiometric amounts, or so called "excess uptake" (Arvin, 1985 and Hogan, 1984).

A great deal of research has been dedicated to the microbiology and the biochemistry of biological phosphorus removal, yet the exact mechanisms involved are still a topic of debate and controversy. Two opposing theories remain, that of biologically mediated chemical precipitation and the theory of biological uptake. Within the latter, three factors are commonly cited as wholly or partially responsible (Buchan, 1983):

1. "Luxury uptake" is the theory which involves phosphorus accumulation due to the cessation of nucleic acid synthesis when the supply of an essential nutrient other than phosphorus or carbon is exhausted. For example, phosphorus may be accumulated as a result of sulfur or nitrogen starvation or due to low pH inhibition.

2. "Overplus Phenomenon" explains excess uptake by postulating that the addition of phosphorus to a system after the organism has been subject to a phosphorus deficient environment induces the organism to uptake more phosphorus than it needs for growth and development.

3. "Population Selection" refers to the selection of populations of microorganisms which are able to out compete other species due to their ability to store polyphosphate under balanced nutrient conditions. Under anaerobic conditions where there are no

exogenous electron acceptors, this stored phosphorus acts as an energy reserve.

Disagreement also centers on the organisms responsible for excess uptake. The genus Acinetobacter is most often mentioned, but Aeromonas and Pseudomonas have also been identified as likely to be responsible for excess phosphorus uptake in some systems (Brodish and Joyner, 1983). In addition, two species of filamentous organisms, Microthrix and Nocardia have been known to accumulate polyphosphate granules in their cells and have been found in a system removing phosphorus down to 0.2 mg/l (Brodish and Joyner, 1983).

In support of biologically mediated chemical precipitation, it has been demonstrated that denitrification in biofilms may lead to calcium phosphate precipitation due to increased pH in the biofilm created by the denitrification reaction (Arvin and Kristensen, 1983). Accumulated phosphate precipitate in the biomass has been found to be as high as 9.3% phosphorus on a dry solids basis (Arvin and Kristensen, 1983).

Perhaps the most widely accepted explanation involves the "population selection" mechanism. It has been suggested that accumulated polyphosphorus in the aerobic section of a system may be used by the organisms as an energy reservoir to sustain them in the anaerobic section which is a part of all biological phosphorus removal systems. This competitive advantage allows the phosphorus removing organisms to thrive by using the stored energy to

accumulate readily biodegradable organics (particularly lower molecular weight fatty acids) in the anaerobic section before other aerobes are able to use them (Marais et al., 1983).

Regardless of the mechanism involved, most researchers agree that several factors are involved in successful operation of a biological phosphorus removal system. Among these are the following (Irvine, 1982):

1. There must be an anaerobic-aerobic staging in the activated sludge system.
2. Orthophosphate is released to the liquid medium in the anaerobic zone.
3. Polyphosphate must be biosynthesized in the aerobic zone and stored in the cells as granules.
4. This phosphorus uptake and release is critical to the survival and proliferation of phosphorus accumulating organisms.

In addition, the accumulation of readily degradable organics by the phosphorus storing bacteria in the anaerobic section seems to be necessary for excess phosphorus uptake. Studies using a Phoredox pilot plant have shown that increasing levels of acetate, propionate and formate enhance phosphorus release while butyrate, hydroxybutyrate, and glucose have relatively little effect on phosphorus release in the anaerobic section (Potgieter and Evans, 1983).



Based on the above discussion, it is apparent that a "generic" biological phosphorus removal system must involve, at a minimum, the following two steps. First, an anaerobic stress period where phosphorus is released and readily degradable organics are sorbed by phosphorus accumulating organisms. Second, an aerobic section where the sorbed organics are oxidized and phosphorus is absorbed in excess of stoichiometric amounts.

2.2 Development Status The Phostrip, A/O and Bardenpho systems were all developed in the late 1960's and 1970's. Phostrip and A/O were developed in the United States by Levin and Air Products and Chemicals Inc., respectively, while the Bardenpho process was developed in South Africa by Barnard.

All three proprietary processes use conventional wastewater treatment equipment which is readily available, and, therefore, equipment procurement and installation should not present reason for delay or expense above those normally encountered in construction. Start-up and operational difficulties have been experienced with each system as would be expected with a developing technology. As a whole, experience to date has been encouraging, and the list of full-scale applications of biological phosphorus removal systems continues to grow.

While each process has been the subject of many pilot plant studies, the Phostrip process has seen the widest plant-scale use in the United States. Many of these plants have been retrofits of

existing facilities, and in some cases the Phostrip process is no longer being used. A need for a more reliable equipment package has been cited, particularly those related to transfer pumps, lime handling facilities, and control instrumentation (Weston, 1984). A listing of full-scale Phostrip plants and their present status is presented as Table 1.

The Bardenpho process has seen little application in the United States to date, but the process has seen impressive application in other countries. Notably in South Africa, where the system was developed in response to an increasing need to recycle water, over thirty wastewater treatment plants are operating or have been designed using the Bardenpho system. A listing of full-scale plants in North America employing the Bardenpho system is presented in Table 2.

The A/O system has been the subject of numerous pilot scale investigations treating a wide variety of wastewaters, but it has only been in the last two years that more plant scale applications have been realized. In addition, many more A/O systems are under construction or being added as a retrofit to existing plants. These plants will certainly bear closer examination as the data base on full scale A/O operation and maintenance grows. A listing of full scale A/O plants is presented in Table 3.

TABLE 1: Status of Phostrip Facilities  
in the United States

Project Location City or Town	Start Up Date	Design Flow M <sup>3</sup> /s	Design Flow mgd	Present Status (Sept. 1986)
Adrian, MI	--	0.31	7.0	data not available
Amherst, NY	1981	1.0	22.0	not in use <sup>1</sup>
Brockton, MA	--	0.80	18.0	data not available
Carpentersville, IL	1979	0.22	5.0	not in use <sup>1</sup>
Ithaca, NY	1987	0.09	2.0	under construction
Landsdale, PA	--	0.11	2.5	not in use <sup>2</sup>
Lititz, PA	--	0.15	3.5	in operation
Reno/Sparks, NV	1974	1.31	30.0	in operation
Rochester, MN	--	0.83	19.0	in operation
Savage, MD	1984	0.66	15.0	in operation
Seneca Falls, NY	1973	0.04	1.0	not in use <sup>3</sup>
Southtowns, NY	1984	0.70	16.0	in partial oper
Tahoe-Truckee, CA	1983	0.32	7.4	in operation
Texas City, TX	--	0.33	7.5	not in use <sup>2,4</sup>

Key: 1. Instrumentation and other mechanical difficulties.  
 2. Permit limits relaxed - system no longer needed.  
 3. Reason for discontinuation unknown.  
 4. Facility loading significantly below design values.

TABLE 2: Current Status of Bardenpho Facilities  
in North America

Project location City or Town	Start Up Date	Design Flow m <sup>3</sup> /s	Flow mgd	Present Status (Sept. 1986)
Palmetto, FL	1979	0.61	1.4	in operation
Kelowna, BC (Canada)	1983	0.26	6.0	in operation
Plukemin, NJ (Hills Development)	1983	0.04	0.85	in operation at reduced flow
Payson, AZ	1984	0.75	1.7	in operation
Fort Meyers, FL (two plants)	1985	1.01	23	in operation
Orange County, FL	1984	0.53 (total)	12	6 mgd in operation 6 mgd in construction
Tarpon Springs, FL	1986	0.18	4.0	under construction

TABLE 3: Current Status of A/O Facilities  
in the United States

Project location City or Town	Start Up Date	Design Flow m <sup>3</sup> /s	Flow mgd	Present Status (Sept. 1986)
Largo, FL	1979	0.14	3.2	in operation
Titusville, FL	1987	0.13	3.0	under construction
Pontiac, MI (E. Boulevard)	1984	0.26 (total)	6.0	3 mgd A/O 3 mgd conv act sldg
Springettsburg, PA	1986	0.66	15.0	under construction
Lancaster, PA	1986	1.31	30	under construction, (retrofit)
Baltimore, MD (Patapsco)	1986	3.07 (4.8 gpm pilot study completed)	70	under construction
Fayetteville, AR	1987	0.48 (pilot plant in testing)	11	under construction
Wayne County, MI (Huron River)	1986	0.53	12	under construction (retrofit)
Rochester, NY	--	0.66 (pilot plant test complete)	15	in design
York, PA	--	1.14	26	in design
Warminstertownship, PA	--	0.35	8.0	in design

2.3 Massachusetts Needs for Nutrient Removal. Within the Commonwealth of Massachusetts, a number of streams or the waterbodies they feed have been identified as sensitive to nutrient loading. In the case of the Ten Mile River, the downstream water is used as a public water supply source in Rhode Island. In the French and Quinnebaug River systems, past algal blooms in dammed and slow moving reaches have made nutrient removal a growing concern. In a number of other streams nutrient removal is a concern largely for the receiving waterbody as in the case of the Taunton and Blackstone rivers which empty into Narragansett Bay.

Presently, there are approximately 150 municipal wastewater treatment plants in operation within the Commonwealth. Twenty seven of these plants have been identified as having effluent phosphorus limits in their operating permits or likely to have such limits in the near future. As the Division of Water Pollution Control of the Massachusetts Department of Environmental Quality Engineering expects very few new plants to be built in Massachusetts in the near future, the potential for the use of biological phosphorus removal lies almost entirely in retrofitting these existing twenty seven plants. Table 4 provides a listing of these plants with their design and average flows, their influent phosphorus and BOD concentrations, their treatment modes and sludge handling systems, and whether or not they have a nitrogen limitation in their operating permits.

The data shown in Table 4 were taken from the years 1979 to 1985

TABLE 4: Massachusetts Treatment Plants with Present  
or Future Phosphorus Removal Needs

Plant Location River Basin	System Type	Des Flow Ave Flow (mgd)	Inf P Range (mg/l)	Inf BOD Range (mg/l)	N Limit	Sludge System
Rockland (North)	AS	2.5 0.85	8.0 4.0-10.0	160 140-200	Y	AnD
Brockton (Tauton)	POx AS	18.0 10.0	15.0 12-18	250 190-370	N	AnD
Mansfield (Taunton)	TF ExA	1.5 0.9	6.0 4.0-8.0	300 110-500	N	GT
Middleborough (Tauton)	AS	2.2 0.8	6.7 3.5-9.0	200 140-450	Y	Ctf
Attleborough (Ten Mile)	AS 2N	6.0 4.5	7.2 5.0-10.0	150 130-220	Y	SDB
N. Attleborough (Ten Mile)	AS 2N	4.61 3.7	8.0 3.0-11.0	180 120-300	Y	Ctf
Medway-CRPCD (Charles)	ExA	4.54 1.8	14.0 4.5-16.0	200 90-350	N	SDB
Milford <sup>1</sup> (Charles)	TF	4.3 2.0	9.5 6.5-13.0	175 150-200	Y	VF
Medfield (Charles)	ExA SF	1.52 0.4	10.0 6.0-16.0	170 100-400	N	CF
Marlborough (East) (SuAsCo)	AS 2N	5.5 2.8	8.0 5.0-12.0	175 50-300	Y	VF
Charlton (French & Quinnebaug)	TF	0.32 0.04	ND	ND	N	ND
Dudley (Fr. & Quinn.)	ExA	0.70 0.80	ND	200 120-300	N	AD
Leicester (Fr. & Quinn.)	ExA	0.118 0.140	ND	ND	N	SDB
Southbridge (Fr. & Quinn.)	AS	2.3 2.8	8.0 6.0-10.0	175 150-300	N	AD
Webster (Fr. & Quinn.)	AS	6.33 3.0	6.0 3.0-9.0	250 150-350	N	SDB

TABLE 4 (continued): Massachusetts Treatment Plants with Present or Future Phosphorus Removal Needs

Plant Location River Basin	System Type	Des Flow Ave Flow	Inf P Range	Inf BOD Range	N Lim	Sludge System
Gardner (Millers)	TF	3.8 2.4	6.5 3.6-8.5	190 150-240	N	AnD SDB
Clinton <sup>1</sup> (Nashua)	TF	6.0 1.8	ND	135 100-175	N	SDB
Leominster (East) (Nashua)	AS 2N	9.3 5.0	6.0 3.0-11.0	180 140-250	Y	VF
Fitchburg (Nashua)	AS 2N	12.5 9.0	10.0 5.0-12.0	200 100-250	Y	DAF VF Inc
Grafton (Blackstone)	AS ExA	3.88 0.7	ND	ND	N	VF
Hopedale (Blackstone)	AS 2N	0.6 0.23	8.2 6.0-11.0	275 220-340	Y	BFP
Worcester (Blackstone)	AS	56.0 30.0	4.5 2.5-7.0	120 80-200	N	BFP Inc
N. Brookfield <sup>1</sup> (Chicopee)	ExA & TF	0.5 0.25	6.0 5.5-6.5	240 125-280	N	ND
Spencer (Chicopee)	AS SF	0.98 0.85	ND	260	N	AD SDB
Palmer (Chicopee)	AS	5.6 3.2	6.5 4.5-7.5	230 180-280	N	CF
Ware (Chicopee)	ExA	2.0 0.5	4.0 3.0-5.5	160 120-210	Y	BFD
Pittsfield (Housatonic)	TF AS 2N	23.0 11.0	5.0 4.0-8.0	240 150-300	Y	AnD SDB

Key: AS= activated sludge, POx=pure oxygen, TF= trickling filter, ExA= extended aeration, 2N= two stage nitrification, AnD= anaerobic digestion, GT= gravity thickeners, SF= sand filtration, Ctf= centrifuge VF= vacuum filter, CF= coil filter, AD= aerobic digestion, SDB= sand drying beds, DAF= dissolved air flotation, Inc= incineration, BFD= belt filter press, ND= no data, Y= nitrogen limit in permit, N= no nitrogen limit in permit,

<sup>1</sup> A new facility is in design or is under construction.



and represent monthly averages which may vary considerably throughout the year. The table is, regardless, useful to illustrate the magnitude of phosphorus removal needs in Massachusetts and the diversity to be found in plants needing phosphorus removal.

Two of the plants have had some experience with biological phosphorus removal systems. The Brockton facility has been upgraded to advanced treatment using the Phostrip process. Unfortunately, it was not possible to visit the facility and information regarding the operation and performance of the plant since its conversion was very limited. The other plant which has some experience with biological phosphorus removal is the North Attleborough Pollution Control Facility which experimented with the A/O process. This plant will be discussed in some detail in chapter four.

2.4 Massachusetts Permit Requirements. In much the same way that facilities vary in their flows, influent characteristics, and receiving waters, so also do their permits. Without ignoring this basic fact, some generalizations can be made. Nutrient limitations in Massachusetts are usually required only during the "summer" months of the year. In practice, these are variously described as April or May through October or November. This aspect of having essentially two different permits for each half year of operation may be an important factor in the consideration of biological systems for the removal of phosphorus. In a plant using a biological phosphorus removal process, it is doubtful that operation would be significantly changed during the year due to the inherent

difficulties associated with the start up and stabilization of microbial populations. If, on the other hand, a plant were using a chemical precipitation process, it may be a relatively simple matter to discontinue the use of the system over the winter.

Another feature of the way that a permit is written which bears consideration is the sampling method and the sampling period to be employed. Phosphorus limitations for some plants are specified as monthly averages while others are specified as not to exceed a maximum concentration of flow weighted composite samples.

A typical permit within the Commonwealth requires an average weekly or monthly phosphorus concentration in the effluent to be less than 1.0 mg/l with a maximum daily concentration of 1.5 or 2.0 mg/l. All three proprietary processes considered in this study have demonstrated the capability of reducing total phosphorus from the 4 to 12 mg/l range typically found in municipal wastewaters down to the 1 to 2 mg/l range (Weston, 1984). The Phostrip process has demonstrated the ability to achieve a 1 mg/l standard at many plants, and Biospherics claims that "In every application, Phostrip use has resulted in effluent containing 0.5 to 1.0 mg/l of total phosphorus while meeting BOD and suspended solids standards" (Biospherics company literature, 1982). In one third of the cases investigated by Weston, however, effluent filters were needed in order to meet these requirements (Weston, 1984). The A/O process has consistently met an average total phosphorus limitation of 2 mg/l at plants in Largo, Florida and Baltimore, Maryland (Deakyné et al., 1983 and Krichten, 1980). In order to consistently meet a 1

mg/l total phosphorus limit an effluent polishing filter to remove particulate phosphorus or a small dose of precipitant prior to the secondary clarifier may be needed. The Bardenpho process has consistently demonstrated average effluent total phosphorus levels of 2.0 to 2.5 mg/l (Barth and Stensel, 1981 and Stensel et al., 1980). It has been further demonstrated that a "minimal" alum dose prior to the secondary clarifier was suitable to produce monthly average total phosphorus levels of approximately 0.5 mg/l (Eimco company literature, 1984).

Another consideration when evaluating the possibility of using biological phosphorus removal processes is whether or not there is a nitrogen limitation in the wastewater treatment plant's operating permit. While Phostrip is applicable principally where only phosphorus removal is required, the process may be used in conjunction with the first stage of a two stage nitrification process. The A/O system may be designed for nitrification or denitrification as well as phosphorus removal, although the ability of the A/O system to provide phosphorus removal and complete nitrification and denitrification remains to be demonstrated (Weston, 1984). The Bardenpho system was originally developed for total nitrogen removal and was subsequently modified to provide phosphorus removal (Hogan, 1984). For this reason the Bardenpho system is generally only used when nitrogen removal is also required.

In Massachusetts, nearly half of the plants needing phosphorus removal also have nitrogen limits in their permits. These limits are generally between 1 and 3 mg/l of ammonia-nitrogen ( $\text{NH}_3\text{-N}$ ).

While nitrification is often needed (where nutrient control is necessary), denitrification is not a requirement within the Commonwealth at the present time. This fact would tend to make the Bardenpho process relatively less attractive when compared to the other process used for nutrient control.

## CHAPTER III

### ALTERNATIVES

3.1 The Phostrip Process. The Phostrip process, by Biospherics Inc., is unique among the three proprietary processes discussed in this study in two respects. First, the Phostrip process is a side stream process. This refers to the fact that the anaerobic section of the process is only seen by a fraction of the plant flow. Generally, 20-30% of the plant flow is taken from the secondary clarifier underflow into the anaerobic phosphorus stripper tank. The remainder of the flow does not experience the anaerobic environment which is central to phosphorus release and excess uptake. The second unique aspect of the Phostrip system is that it is, by design, a combination biological-chemical system. Lime is used to precipitate the phosphorus which is released from the stripper tank when the activated sludge microorganisms are subjected to the stripper tank's anaerobic environment.

The Phostrip process is depicted schematically in Figure 1. The major treatment units in the system are an activated sludge process consisting of a primary clarifier, aeration basin, and secondary clarifier and the phosphorus stripping section consisting of an anaerobic phosphorus stripping tank, a lime feed system, and a lime mixing tank or lime reactor/clarifier. If a lime mixing tank is used instead of a reactor/clarifier, the resulting calcium phosphate is precipitated in the primary clarifier.

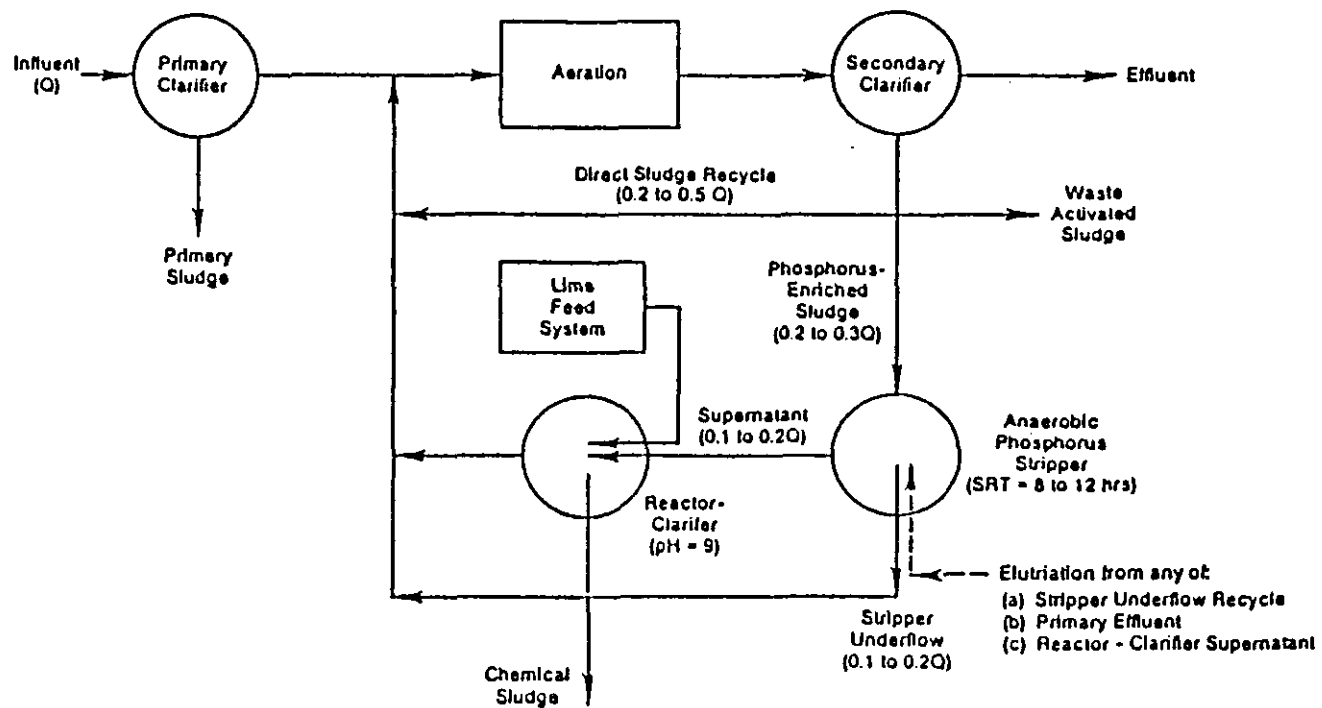


Figure 1. Phostrip Process Flow Diagram.

According to Biospherics company literature, the process is suited to use with conventional, tapered, step modified and pure oxygen aeration systems (Biospherics, 1985). The process may not be applicable with some contact stabilization and extended aeration systems as the Phostrip process has a hydraulic detention time limitation of between 1 and 10 hours in the aeration basin. The system is able to operate in conjunction with systems designed for nitrification. Modifications may be required to either exclude nitrate-nitrogen from the stripper tank or to increase the detention time in the stripper tank in order to achieve the needed phosphorus release (Weston, 1984).

In the process, flow from the secondary clarifier is divided into four streams: (1) the clarifier overflow (this is the plant effluent), (2) the direct return sludge, (3) the waste sludge, and (4) the sludge to the stripper tank. As long as the sludge blanket in the secondary clarifier is maintained in an aerobic condition, the sludge drawn from the clarifier will be rich in phosphorus. Phosphorus is removed from the system in two ways. The wasted sludge accounts for one portion of the phosphorus removal while the sludge sent to the stripper tank represents the other means of phosphorus removal.

The sludge sent to the stripper tank normally represents 20 to 30% of the influent flow depending on the extent of phosphorus removal needed. The hydraulic detention time in the anaerobic stripper tank is generally between 8 and 12 hours, in which time

phosphorus is released to the surrounding liquid. During this time a portion of the active biomass undergoes lysis which provides the readily biodegradable organic matter required to achieve phosphorus release (Arvin, 1985). The liquid which is now rich in phosphorus is then elutriated from the stressed organisms by flow from one of the following sources: (1) stripper tank underflow recycle, (2) primary clarifier overflow, (3) secondary clarifier overflow, or (4) supernatant from the reactor/clarifier (if used). The choice of elutriant may be based upon plant limitations or retrofit requirements, but Arvin has suggested that primary effluent may be the most effective elutriant as this source contains the largest quantity of soluble BOD to enhance phosphorus release (Arvin, 1985).

Phosphorus rich supernatant from the stripper tank is then precipitated with lime. Because the supernatant represents only one-tenth to one-fifth of the plant flow and calcium phosphate precipitation is pH dependent (and independent of the amount of phosphorus precipitated due to many side reactions), chemical costs are only 10 to 20% of those encountered in conventional precipitation. The phosphorus starved stripper underflow is returned to the aeration basin with the direct recycle sludge where excess uptake occurs and the process is repeated.

According to Hogan (1984), The use of a reactor/clarifier instead of a mixer with precipitation in the primary clarifier has four basic advantages: (1) This allows the chemical sludge to be handled independently, (2) It is easier to maintain the pH near the optimal



range of 9.0 for calcium phosphate precipitation, (3) One has the ability to recover lime from the reactor/clarifier sludge, and (4) Keeping the primary sludge separate from the calcium phosphate sludge avoids problems of more difficult dewatering encountered with mixed primary and chemical sludges.

The process has achieved effluent phosphorus concentrations of less than 1 mg/l under a wide range of climatic conditions. A brief listing of some major wastewater characteristics and design parameters based on Biospheric's past experience with the process is shown in Table 5.

Major factors which influence the degree of phosphorus removal obtained from the Phostrip system include (Peirano, 1977):

1. The volume of sludge taken from the clarifier into the stripper tank,
2. the solids detention time in the stripper tank, and
3. the rate of elutriation from the stripper tank.

Among the advantages which have been cited for the Phostrip process over conventional chemical precipitation and other biological processes are these (Biospherics, 1985 and Weston, 1984):

1. The system can operate over a wide range of influent BOD, phosphorus concentrations and aeration times,
2. the stripper tank can hold a reserve of healthy microbes to shield the system from shock loads and toxic materials,

TABLE 5: Phostrip design parameters and typical  
wastewater characteristics

Parameter	Parameter Value
Inf. BOD	70 to 300 mg/l
Inf. Phosphorus	3 to 20 mg/l
Temperature	10° to 30°C
NO <sub>2</sub> and NO <sub>3</sub> -nitrogen	1 to 30 mg/l
Aeration tank HDT	1 to 10 hours
MLSS	600 to 5,000 mg/l
Stripper SRT	8 to 12 hours
Sidestream Sludge Flow	20 to 30% of inf.
Stripper Supernatant Flow	10 to 20% of inf.
Elutriant Flow	50 to 100% of str. feed
Lime Dosage	100 to 300 mg/l

3. the system has been able to consistently produce an effluent phosphorus concentration below 1 mg/l total phosphorus.

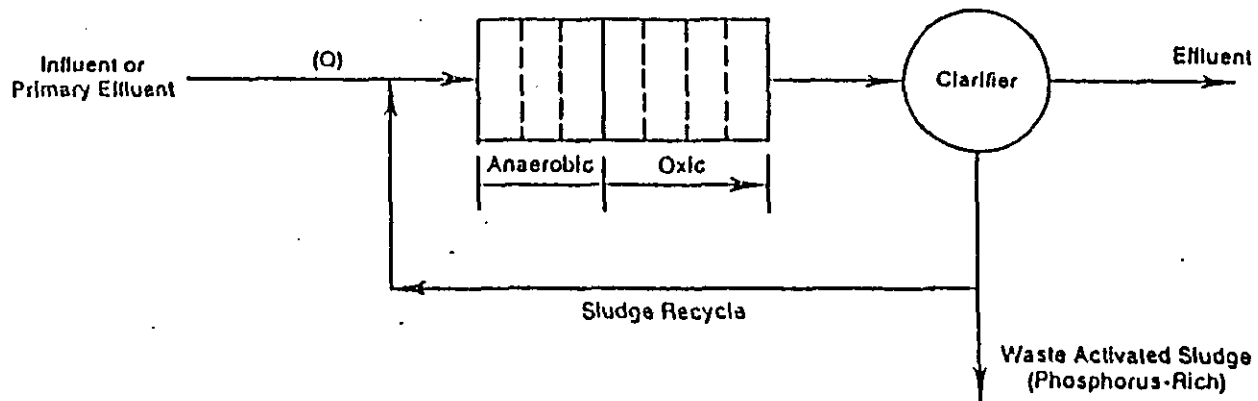
4. the system uses a fraction of the chemicals required for conventional chemical precipitation,

5. there is no leakage of chemical precipitants in the effluent from the system which might be harmful to a receiving waterbody, and

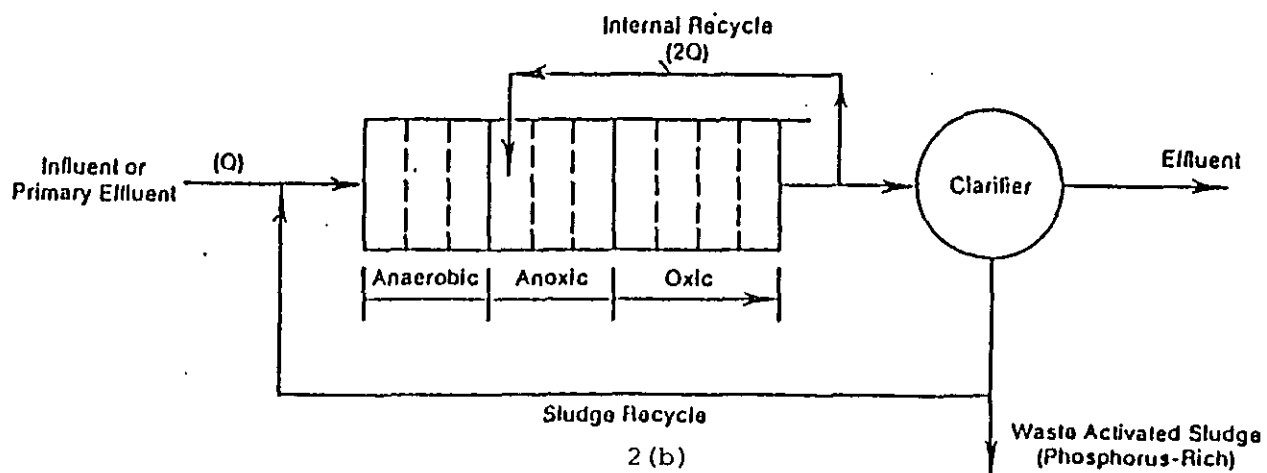
6. phostrip requires less tankage than other biological phosphorus removal systems, as only a portion of the plant flow is held in an anaerobic environment.

3.2 The A/O Process. The A/O process, by Air Products and Chemicals, is a mainstream suspended growth process which combines a staged anaerobic section with a conventional aerated sludge system. The process produces a single sludge which is the waste activated sludge from the secondary clarifier. The process may be designed for phosphorus removal only, phosphorus removal with nitrification, or phosphorus removal with nitrification and denitrification. Initial experience with A/O design and operation for phosphorus removal and nitrification/denitrification at Largo, Florida showed that both phosphorus removal and complete denitrification were not achieved consistently (Weston, 1984). Figure 2(a) is a schematic representation of the A/O process for phosphorus removal only while Figure 2(b) shows the process as designed for phosphorus removal with nitrification/denitrification.

When denitrification is not required, the process consists of anaerobic and aerobic sections which are partitioned into several completely mixed stages to prevent backmixing and to approach plug flow. The anaerobic section is deficient in both dissolved oxygen and chemically bound oxygen. Influent wastewater, with or without primary clarification, is mixed with the recycle sludge from the secondary clarifier at the inlet to the anaerobic section.



2 (a)



2 (b)

Figure 2. A/O Process Flow Diagrams.

This section is covered to exclude oxygen and is completely mixed. The purpose of this section is to induce the stress condition which makes the microorganisms release phosphorus and absorb readily biodegradable organics from the influent. In the aerated section, the sorbed organics are oxidized to provide energy for cell growth and division. A small fraction of the energy liberated in the oxidation reaction is used to accomplish the excess phosphorus uptake. Phosphorus is removed from the system through wasting a portion of the phosphorus rich biomass after settling in the secondary clarifier. This waste sludge typically contains 4.2 to 6.0% phosphorus by dry weight (Hong et al., 1982). In this mode of operation the A/O process is a high rate system with a detention time in the aerobic section between 1 and 3 hours and an overall hydraulic retention time of between 1.8 and 3.5 hours (Air Products and Chemicals, 1984).

When nitrification is required, the retention time in the aerobic section is increased to between 2 and 5 hours with an overall retention time between 2.5 and 5.5 hours. The mixed liquor suspended solids concentration is also increased from the 2,000 to 4,000 mg/l range typical of the system when used for phosphorus and BOD removal only to between 3,000 and 5,000 mg/l.

When denitrification is required, an anoxic section is added between the anaerobic and aerobic sections. In this section there is no dissolved oxygen but chemically bound oxygen is present in the form of nitrates and nitrites. In this configuration, nitrified

mixed liquor from the last aerobic section is recycled to the first anoxic stage. This recycle flow is generally twice the plant flow. Air Products states that there is a sufficient level of BOD from the preceding anaerobic section to serve as an energy source for the denitrification so that external energy sources such as methanol are not required (Hong et al., 1982).

Air Products has developed a set of design and operating parameters based on their experience with laboratory studies, pilot scale plants, and their full scale experience at Largo, Florida. These parameters are shown in Table 6 (Air Products and Chemicals, 1984).

The most critical factor in maintaining a low phosphorus effluent with the A/O system is the influent soluble BOD to phosphorus ratio (Hong et al., 1982). This ratio should be 10 or greater if effluent levels of 1.0 mg/l are to be expected. This is normally the case for municipal wastewaters in the United States.

Phosphorus removal is also dependent upon several other factors (Hogan, 1984):

1. Anaerobic conditions in the first stage must be maintained so that there are no exogenous electron acceptors,
2. clarifier overflow must be low enough to prevent solids from being discharged with the effluent or polishing filters will be needed,

TABLE 6: A/O Design and Operating Parameters

Parameter	BOD, P Removal	BOD, P, N Removal
Detention time (hr)		
Anaerobic	0.5-1.0	0.5-1.0
Anoxic	not req'd	0.5-1.0
Oxic (Aerobic)	1.0-3.0	2.0-5.0
Food to microorganism ratio F/M (mg/l / mg/l)	0.2-0.6	0.15-0.25
Mixed Liquor Suspended Solids Concentration MLSS (mg/l)	2000-4000	3000-5000
Oxygen Use (kg/kg BOD)	1.0	1.2
Return Sludge Flow (% of inf)	10-30	20-50
Underflow Conc. (% solids)	2-4	1.5-3.0
Internal Recycle Flow (% inf)	not req'd	100-300
Minimum D.O. Aerobic (mg/l)	2.0	2.0
Waste Sludge (kg/kg BOD rem)	0.5-0.8	0.3-0.6
Mixing Energy Anaerobic Section (KW/1000 liters)	0.15	0.15

3. The secondary clarifier must be kept aerobic to ensure that phosphorus bleedback will not occur,
4. Return flow from sludge treatment must be restricted so that excessive phosphorus loads are not returned to the wastewater flow train, and
5. The soluble food to microorganism ratio should be above 0.08.

Among the advantages which have been cited for the A/O process over conventional precipitation and other biological processes are the following (Air Products and Chemicals, 1984 and Hogan, 1984):

1. There is no chemical cost and no need for chemical storage and handling,
2. The high rate system has low tankage requirements,
3. No chemical sludges requiring special treatment, handling or disposal are generated,
4. The system is energy and cost efficient. Mixing and aeration requirements are minimal and less return sludge needs to be pumped,
5. The resulting sludge has higher nutrient levels, thereby increasing its value for composting and final sale as a soil conditioner,
6. The process lends itself to low capital, simple retrofitting of existing activated sludge systems, and



7. operation of the A/O system is simple and very similar to standard operation of a conventional activated sludge system.

3.3 The Bardenpho Process. The Bardenpho process, marketed by Eimco Process Equipment Co., is designed for phosphorus and BOD removal with complete nitrification and denitrification. The name itself is an acronym derived from the words Barnard, denitrification and phosphorus. The system is generally only used when nitrogen removal is required. In fact, the system was first designed to accomplish nitrification and denitrification only. Phosphorus removal was added to the system latter (Hogan, 1984). The process is similar to the A/O system with denitirification which was discussed earlier in that it is a single sludge suspended growth system consisting of anaerobic, aerobic and anoxic treatment zones.

The Bardenpho system is shown schematically in Figure 3. In the Bardenpho system five treatment zones are provided, one anaerobic zone, two anoxic zones, and two aerated (or oxic) zones. The influent to the system may be primary clarifier effluent or raw wastewater depending on the organic strength of the wastewater.

The first stage is the anaerobic or fermentation stage. In this tank, the influent is mixed with the return sludge from the secondary clarifier. The tank is covered to exclude oxygen. Any dissolved or chemically bound oxygen which enters the tank with one of the streams is quickly used by the activated sludge organisms so that there are no exogenous electron acceptors. As in the A/O

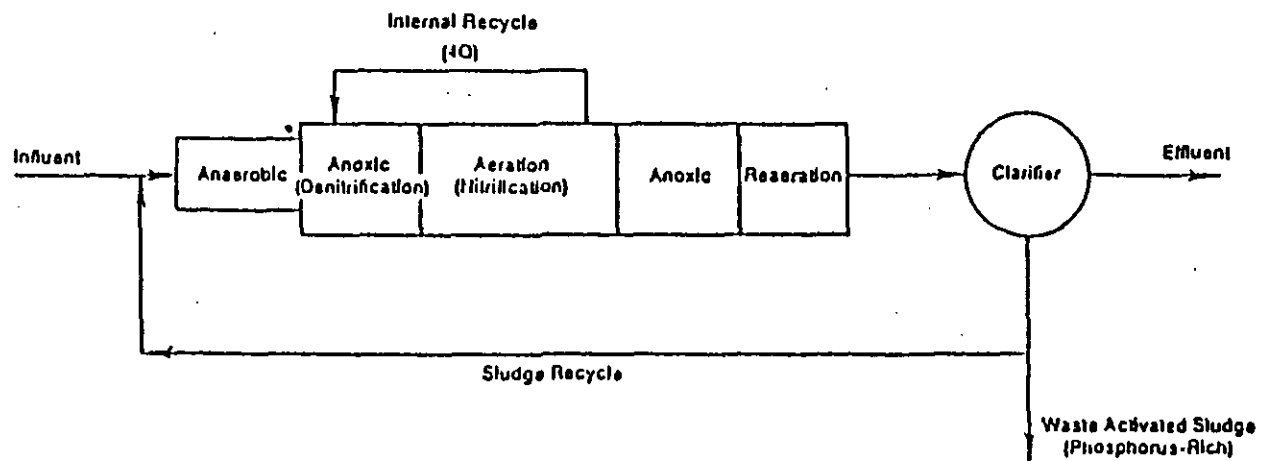


Figure 3. Bardenpho Process Flow Diagram.

process, the phosphorus storing organisms absorb much of the readily biodegradable organics in the influent and simultaneously release phosphorus to the surrounding liquid. At the Palmetto, Florida facility, grab samples showed ortho-phosphorus levels in the range of 14-20 mg/l, or at least three times the plant influent concentration. Stage one detention times are typically 0.6 to 2.0 hours (Weston, 1984).

The second tank is the first anoxic stage. Mixed liquor from the aerated third stage is mixed with the flow from the fermentation zone. This internal recycle is typically four to five times the plant influent flow (Stensel et al., 1980). Because the tank is covered, there is no dissolved oxygen but chemically bound oxygen from the third stage is available. The treatment objective in this stage is to reduce the nitrates from the third stage to gaseous nitrogen (denitrification). Eimco states that around two-thirds of the influent nitrogen is removed in this tank (Eimco Process Equipment Company, 1984). The BOD from the fermentation zone acts as a food source so that no methanol or other chemical additions are required. Retention times in this section are typically 2.2 to 5.2 hours (Weston, 1984).

The third tank is the first aerated section or nitrification zone. This step is essentially an extended aeration biological treatment step (Burdick and Dallaire, 1978). The treatment objectives here are BOD reduction, conversion of ammonia to nitrate (nitrification), and excess phosphorus uptake. The nitrification

stage is designed on the basis of providing a sufficient solids retention time as a function of temperature for nitrification and, sludge stabilization if so desired (Stensel et al., 1980). Detention times for this stage are typically between 6.5 and 19.0 hours with the average detention time near 11 hours (Weston, 1984).

The fourth tank is a second anoxic stage which acts as a polishing step to convert any remaining nitrates to nitrogen gas. Because the available BOD is low, some bacteria undergo lysis and thus provide additional food for the remaining active bacteria. Due to food limitations and the time required for lysis and subsequent nutrient extraction, the reaction is slow. Phosphorus taken up in the third tank remains in the active biomass due to the presence of chemically bound oxygen in the nitrates (Burdick and Dallaire, 1978). Typical detention times for this stage range between 2.2 and 5.7 hours (Weston, 1984).

In the fifth, or reaeration stage, any phosphorus released by lysed bacteria in the second anoxic stage should be taken up by the active phosphorus accumulating organisms and denitrification is stopped. It is important to stop the denitrification reaction so that escaping nitrogen gas does not hinder sludge settling in the secondary clarifier. Another important function of this stage is to increase the dissolved oxygen content of the wastewater so that the sludge blanket in the secondary clarifier will not go anaerobic and release its accumulated phosphorus. Detention time in the reaeration stage averages one hour.

Typical design and operating parameters which have been developed from Bardenpho experience in South Africa and the United States are shown in Table 7 (Hogan, 1984 and Barnard, 1983).

TABLE 7: Bardenpho Design and Operating Parameters

Parameter	Parameter Value
Solids Retention Time (days)	15 to 40
MLSS (mg/l)	3,500 to 5,000
Temperature	above 10°C
Third Stage of Dissolved Oxygen (mg/l)	1 to 2.0 mg/l
Internal Recycle Rate (% inf.)	400 to 500
Inf. BOD to Phosphorus Ratio	25:1
Sludge Recycle Ratio (%inf.)	100
Sludge Blanket Depth (m.)	0.15 to 0.45

Many additional factors have been identified as important to the successful operation of a Bardenpho system (Barnard, 1983). These include:

1. The COD:TKN (Chemical oxygen demand to total Kjeldahl nitrogen ratio) should be near ten or above,
2. phosphorus and nitrogen removal increase with increasing pH,
3. flexibility of dissolved oxygen control in the third stage is very important as too little oxygen will result in poor

phosphorus uptake and too much dissolved oxygen will inhibit denitrification in stages 2 and 4,

4. the process functions better at solids residence times in excess of 30 days while producing a more stable sludge, and

5. the secondary clarifier should be designed conservatively so that solids rich in phosphorus will not be contained in the plant effluent.

Advantages which have been cited for the Bardenpho system over chemical precipitation and other biological processes include (Eimco Process Equipment Company, 1984 and Hogan, 1984):

1. No chemicals are required so that there is no need for chemical handling and no chemical sludge to be disposed,

2. simple design minimizes construction costs,

3. the process is simple to operate being quite similar to a conventional extended aeration system,

4. long solids retention times lend themselves to process stability and mitigate shock loadings,

5. the stable sludge which is produced may not need additional stabilization, thus allowing direct disposal,

6. little alkalinity is destroyed in the process due to the system's extensive denitrification,

7. separation of treatment units lends itself to finer system control capability, and

8. the phosphorus content of the sludge makes it more attractive as a soil conditioner.

3.4 Chemical Precipitation Systems. Phosphorus removal in the past has most often been accomplished through the use of ionic forms of aluminum, iron, or calcium as precipitants. This is also the most widely used method in the Commonwealth today. Precipitants which have been used successfully in the past include hydrated aluminum sulfate "alum" ( $\text{Al}_2(\text{SO}_4)_3 \cdot 14\text{H}_2\text{O}$ ), sodium aluminate ( $\text{Na}_2\text{Al}_2\text{O}_4$ ), ferrous sulfate ( $\text{Fe}(\text{SO}_4)_2 \cdot 7\text{H}_2\text{O}$ ), ferric chloride ( $\text{FeCl}_3$ ), ferrous chloride "waste pickle liquor" ( $\text{FeCl}_2$ ), and lime ( $\text{Ca}(\text{OH})_2$  or  $\text{CaO}$ ) (U.S. EPA, 1976). In a conventional activated sludge plant, the precipitant may be added before the primary clarifier, into the aeration basin or before the secondary clarifier depending on the treatment objectives and available facilities. In some cases, chemical precipitation may be accomplished in a separate tertiary system.

In order to practice chemical precipitation, additional facilities for chemical storage and feeding and sludge handling may be required. In addition, the plant operator must be able to accurately meter the plant influent and determine changes in the influent wastewater composition quickly as chemical feed rates may need adjustment. A simple generalized flowsheet for chemical precipitation is shown in Figure 4.

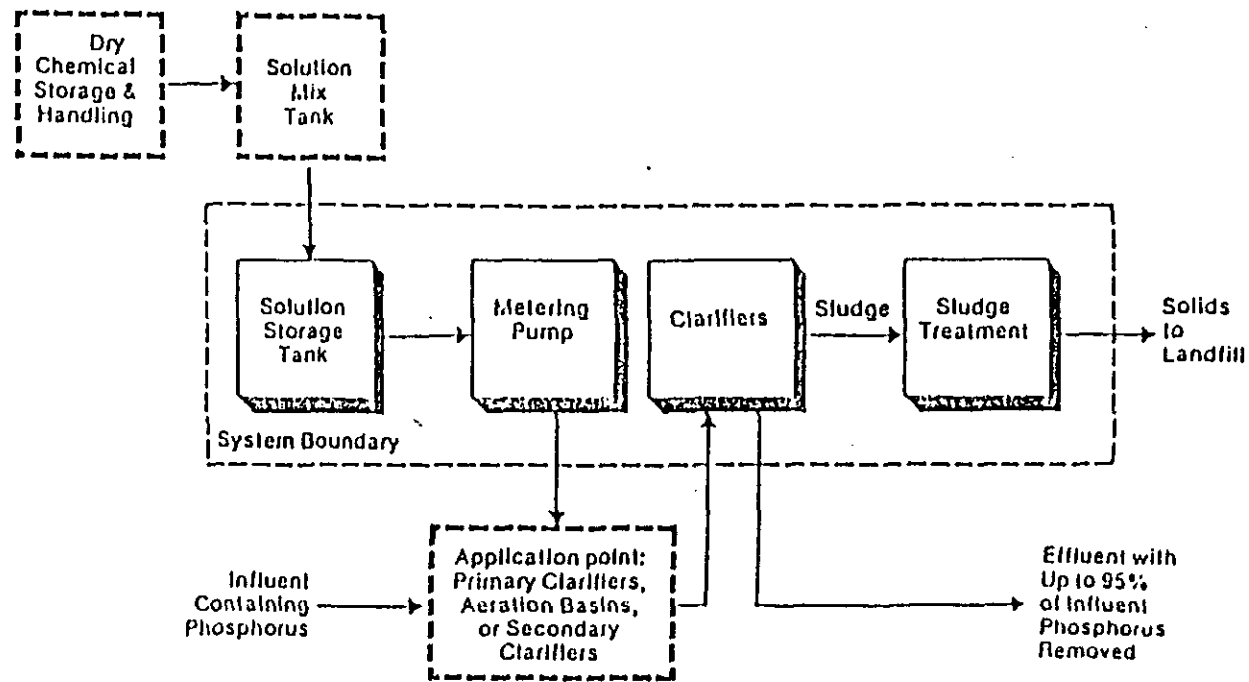


Figure 4. Phosphorus Removal by Chemical Addition.



An important factor in the precipitation of phosphorus from wastewater is pH. Aluminum phosphate compounds formed with the use of alum or sodium aluminate exhibit a minimum solubility around pH 6.3. Ferric phosphate formed with the use of iron salts is least soluble at a pH of 5.3 (Metcalf and Eddy, 1979). Calcium phosphate precipitation using lime may be practiced at pH levels of 9.5 (low lime treatment) to 11.0 (high lime treatment) (U.S. EPA, 1976). Because treatment systems in practice are often operated at pH values which are not optimum, greater chemical dosages may be required. Lime and pickle liquor use is limited because they produce low phosphorus effluent only at high pH levels (Metcalf and Eddy, 1979). When the wastewater is low in alkalinity, sodium aluminate may be preferred as alum and ferric chloride cause a reduction in alkalinity while aluminate raises the alkalinity (U.S. EPA, 1976). In some systems lime is added with ferric chloride or alum in order to maintain the alkalinity of the system. This possible loss of alkalinity is especially important in nitrifying plants as nitrification also destroys alkalinity.

Mineral salts are generally applied in the range of 1 to 3 moles of metal ions for each mole of phosphorus to be removed (Metcalf and Eddy, 1979). The exact rate of application should be determined from on site testing as dosage varies considerably with wastewater characteristics, effluent requirements, chemical purity and point of addition.

The Environmental Protection Agency has estimated "typical" alum additions as a function of percentage of phosphorus removal needed. For an 85% reduction, an alum to phosphorus weight ratio of approximately 16:1 is needed while 95% reduction requires a ratio of 22:1 (U.S. EPA, 1976). For example, in order to achieve a 1 mg/l phosphorus effluent, when the influent contains 8 mg/l (this value being representative of plants shown in Table 4), then a reduction of slightly more than 85% is required. The alum dosage would then be approximately,

$$16(8) = 130 \text{ mg/l} = 1070 \text{ lbs/million gallons}$$

Theoretical dosages may be similarly determined for ferric chloride and sodium aluminate, but actual dosages will vary from the theoretical values and should be determined through on-site testing.

If lime addition is to be practiced, the amount of lime added to the wastewater must be sufficient to combine with all the free carbonic acid and calcium bicarbonate in the wastewater so that the excess calcium ions can react with the ortho-phosphate to form insoluble hydroxyapatite (Hogan, 1984). Recarbonation will often be necessary after lime treatment in order to lower the pH and prevent scaling (Metcalf and Eddy, 1979).

As a rough indication of the relative dosages required for the use of different precipitants, the EPA has estimated that in order to reduce an influent phosphorus of about 10 mg/l to less than 3.0

mg/l, 200 mg/l of alum, 100 mg/l ferric choride or 150 mg/l of lime would be required (U.S. EPA, 1984).

The choice of the point of the precipitant addition is governed by several factors such as plant capacity and loading, sludge handling abilities, and the costs of sludge disposal. Table 8 lists the relative advantages and disadvantages of various points of chemical addition (Hogan, 1984 and Metcalf and Eddy, 1979).

Chemical costs are normally the largest fraction of the cost incurred in using chemical precipitation. Costs are influenced by the type of chemical used, power requirements, sludge disposal methods and possible chemical reuse. In one study, chemical and operating costs accounted for 80% of the total phosphorus removal cost. In the case of iron and aluminum salts, 70% of this was chemical cost (Hogan, 1984).

Phosphorus removal by chemical addition will increase the amount of sludge produced in a plant and often makes the sludge more difficult to dewater and dispose. Chemical addition has increased sludge volumes by as much as 60% and sludge mass by 40% (Hogan, 1984). In a study which included four plants removing phosphorus with chemical precipitation, Switzenbaum et al. (1981) found an increase in mass of between 2.5 and 19.6%. Other studies have shown increases in sludge volume to be typically in the range of 10 to 25% (Hogan, 1984 and U.S. EPA, 1976).

If chemical addition is chosen for phosphorus removal, the choice of chemical to be used should be based upon several factors (Metcalf and Eddy, 1979):

TABLE 8: Advantages and Disadvantages of  
Chemical Addition at Different  
Points for Phosphorus Removal

Point of Addition	Advantages	Disadvantages
Primary Clarifier	Applicable to most plants; significant BOD and suspended removal; may reduce aeration basin loading; lowest degree of metal leakage	Least efficient sludge is more difficult to dewater; will not precipitate phosphorus not yet converted to ortho-phosphate
Aeration Basin	Lowest cost; lower chemical dosage than primary; improved stability of waste sludge; recycle of sludge provides a precipitant reservoir	Metal overdose may cause low pH toxicity; cannot use lime due to high pH requirements increased sludge recycle due to inerts
Secondary Clarifier	Lowest effluent phosphorus; lime recovery is possible; most efficient use of chemical	Highest capital cost; highest metal leakage

1. Influent phosphorus level,
2. wastewater suspended solids and alkalinity,
3. chemical costs,
4. reliability of chemical supply,
5. sludge handling facilities,
6. ultimate disposal methods,
7. compatibility with other treatment processes in use in the plant, and
8. potential adverse environmental effects of the chemical used.

Some of the advantages of chemical precipitation over biological methods of phosphorus removal are these:

1. The method is reliable and well understood,
2. the method requires little capital investment,
3. chemicals may be changed and adjusted freely to match changes in flow and wastewater composition,
4. chemical feed systems may be shut down and started up easily to meet seasonal requirements, and
5. simple retrofit requirements make chemical addition applicable to nearly any plant or treatment mode.

## CHAPTER IV

## CASE STUDIES

In the course of this research, three existing municipal wastewater treatment plants having phosphorus limitations in their present effluent permits were selected for further study. The selected plants were chosen on the merits of their ability to represent the Commonwealth as a whole in terms of geography (east to west), size (in terms of plant flow), industrial contribution, and effluent permit limitations. In addition, the plants studied have been in general compliance with their permits so that valid comparisons could be made between present practice and possible retrofit to biological phosphorus removal systems.

4.1 Pittsfield Wastewater Treatment Plant. The Pittsfield Wastewater Treatment Plant was expanded in 1963 and again in 1974 under the design of Camp Dresser and McKee Inc. to treat  $0.75 \text{ m}^3/\text{s}$  (17 mgd) of domestic and industrial wastewater from the communities of Pittsfield, Dalton, Lanesborough and North Lenox. The present facility is expected to meet the area's needs through 1995 and is currently treating a summertime flow of approximately  $0.44 \text{ m}^3/\text{s}$  (10 mgd). A schematic representation of the plant is shown in Figure 5. The plant discharges to the Housatonic River which is presently designated a "class C" waterbody.

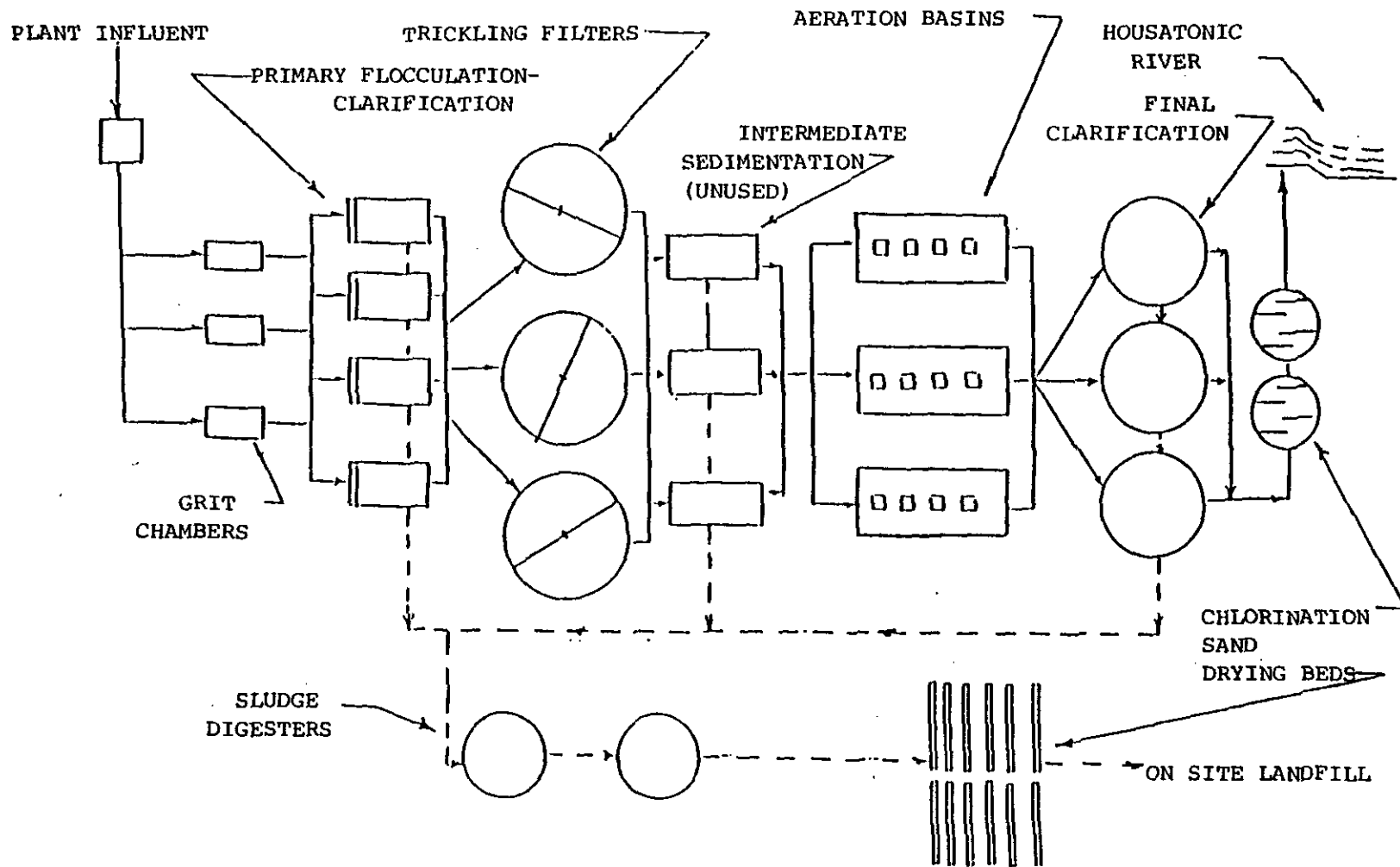


Figure 5. Pittsfield Wastewater Treatment Plant Schematic Diagram.

Influent enters through a 183 cm (72 inch) gravity sewer. Preliminary treatment facilities consist of magnetic flow metering, bar racks and grit chambers. Three  $0.65 \text{ m}^3/\text{s}$  (15 mgd) vertical lift pumps send the flow to the plant's primary clarifiers.

Primary settling is achieved with four rectangular settling basins, each of which is preceded by a flocculation basin equipped with a low speed vertical mixer. Flocculent is not added to the flow at this point. Waste activated sludge from the secondary clarifier is added to the primary basins in order to obtain a higher solids concentration in the secondary sludge. The overflow rate is  $16.25 \text{ m}^3/\text{m}^2/\text{day}$  ( $399 \text{ gpd}/\text{ft}^2$ ) at design flow. The concentration of the sludge leaving the primary clarifier is typically 3 to 4% solids by weight.

The primary effluent is then pumped to high rate trickling filters which function as the first stage of the plant's two stage nitrification system. There are three rock media filters 61 meters (200 feet) in diameter and 1.8 m (6 ft) deep, only two of which are presently being used. A recycle capability is included to help maintain a desirable hydraulic loading. Recycle is not normally used at present, however. The trickling filters were designed to operate at hydraulic loadings between 125 and 630 l/min (2,000 and 10,000 gpm) with a maximum fluid velocity of 1.19 m/s (3.9 fps).

Three humus settling tanks were provided in the last expansion, but these have not been used and the flow is by-passed to the aeration basins. These "intermediate clarifiers" or humus settling



tanks were designed to settle the biofilm which sloughs off the rock media of the trickling filter. The plant operators feel that using these clarifiers would reduce the BOD load to the second stage too much to accomplish good nitrification.

The second, or nitrifying stage, is made up of three rectangular aeration basins. Each basin is equipped with four surface aerators rated at 29.8 KW (40 hp) each. Generally, only two or three of the aerators in any one basin are used. Dissolved oxygen in the aeration basins is targeted to be between 2 and 5 mg/l and averages around 3 mg/l. At design flow, the hydraulic detention time in the basins is 6.56 hours and the mixed liquor suspended solids are maintained between 1500 and 2500 mg/l.

Three circular sedimentation tanks serve as secondary clarifiers. At design flow, the detention time is 6.69 hours and the overflow rate is  $16.5 \text{ m}^3/\text{m}^2/\text{d}$  ( $405 \text{ gpd}/\text{ft}^2$ ). The clarifier was designed to have a sludge blanket .76 meters (2.5 ft) deep, but the sludge blanket is often 1.2 to 1.5 meters (4 to 5 ft). Between 50 and 80% of the solids are recycled to the aeration basin while the remaining waste activated sludge goes to the primary sedimentation basin. The sludge leaving the secondary clarifier generally contains 1% solids by weight. During the summer, the effluent is chlorinated prior to discharge to the Housatonic River.

Combined primary and secondary sludge is digested anaerobically in a two stage system. The digesters yield approximately  $2,000 \text{ m}^3$  ( $70,000 \text{ ft}^3$ ) of 65% methane gas per day which is burned in two

dual fuel engines to produce a portion of the power required for wastewater and sludge pumping. Engine cooling water is used to heat the digesters.

Following digestion, the sludge, which is now approximately 2% solids, is sent to 4.5 hectare (11 acres) of sand drying beds before being landfilled on site. It is expected that dewatering by belt presses will be added in the future and the area devoted to sand beds will be used for landfilling sludge.

The plant is required to both remove phosphorus and convert ammonia nitrogen (nitrification) during the summer months (May 15 to September 15). BOD and suspended solids limits had been lower in the summer period but this is not the case with the present permit. Daily and weekly permit limitations and average influent characteristics for the Pittsfield facility are shown in Table 9.

Phosphorus removal is currently accomplished by chemical addition with sodium aluminate, although alum has been used in the past. Aluminate has proven to be helpful in maintaining the alkalinity necessary for nitrification in the aeration basin and helped the plant operators avoid low pH inhibition in the anaerobic digesters. An aluminate solution is added at the aeration basin overflow immediately prior to the secondary clarifiers. Plant operators believe the chemical precipitation operation causes an increase in sludge volume of 15 to 20% which is nearly double the increase which had been predicted. Phosphorus removal without

TABLE 9: Pittsfield Permit and Influent Values

Parameter	Influent (mg/l)	Permit Limit	
		max day (mg/l)	avg week (mg/l)
BOD <sub>5</sub> (summer)	280	15.0	10.0
BOD <sub>5</sub> (winter)	200	15.0	10.0
Susp. Solids (sum)	230	30.0	20.0
Susp. Solids (win)	170	30.0	20.0
Ammonia - N (sum)	15.0	1.5	1.0
Phosphorus (sum)	6.0	1.5	1.0
Flow (win-spr)	.57 m <sup>3</sup> /s (13 mgd)	1.0 m <sup>3</sup> /s (23 mgd)	
Flow (sum-fall)	.44 m <sup>3</sup> /s (10 mgd)	1.0 m <sup>3</sup> /s (23 mgd)	

chemical addition is estimated to be approximately 45% by normal sedimentation and biological uptake, while 85% removal is accomplished with the present sodium aluminate system.

The plant is running well overall, and effluent levels are consistently under permit limits. The large volume of sludge produced in the secondary clarifiers has been a continuing problem

which could be solved with the addition of a sludge thickening system. Typically, the secondary sludge has a sludge volume index of 135. A second problem involves the load and alkalinity fluctuations attributed to periodic shutdown of area paper mills. Particularly when alum was used for phosphorus removal, plant performance would deteriorate coincident with summer shutdown and holiday periods. Plant operators feel that papermill contributions boost the influent alkalinity by approximately 75 mg/l. A third problem which is encountered irregularly is foaming in the effluent during summertime chlorination.

Biological phosphorus removal is probably not applicable to this plant for the future for a number of reasons. First, the use of trickling filters for the first stage of the two stage nitrification process could prove to be problematic. Though some research is underway, the technology of biological phosphorus removal with fixed film systems is not yet available. Second, the sludge handling problems presently being encountered would have to be solved before a biological phosphorus removal system could be operated. Sludge blanket depths of 1.2 meters (4 ft) would be difficult to keep aerobic. A sludge blanket under two feet is often recommended in order to prevent the anaerobic conditions which cause phosphorus release and subsequent washout with the plant effluent (Weston, 1984). Third, anaerobic digestion of the sludge would cause phosphorus release. The phosphorus would then require chemical precipitation. While it is quite possible that chemical costs may

be significantly reduced by using less chemical or a less expensive chemical to precipitate this small sidestream, it is doubtful that these savings would be able to justify the installation of an entire biological phosphorus removal system in Pittsfield.

4.2 Palmer Wastewater Treatment Facility. The Palmer Wastewater Treatment Facility was designed by Tighe and Bond/SCI Consulting Engineers to treat an average flow of  $0.25 \text{ m}^3/\text{s}$  (5.6 mgd). Current flow averages between  $0.13$  and  $0.17 \text{ m}^3/\text{s}$  (3 and 4 mgd). The plant was dedicated and began operation in 1980 treating domestic and industrial wastewater from the villages of Thorndike, Bondsville, Three Rivers and the town of Monson as well as Palmer itself. The plant effluent is discharged to the Chicopee River. (The plant is represented by the flowsheet which is Figure 6.)

Wastewater enters the plant through a 91.5 cm (36 in) sewer and a 30.5 cm (12 in) interceptor. Preliminary treatment consists of a mechanically cleaned bar rack, an aerated grit chamber, and two comminutors. Wastewater then flows by gravity to the primary clarifiers.

Two circular tanks 23 meters (75 ft) in diameter and 2.5 meters (8 ft) deep serve as primary clarifiers. The overflow rate at design flow is  $25.8 \text{ m}^3/\text{m}^2/\text{d}$  ( $634 \text{ gpd}/\text{ft}^2$ ) and the detention time is 2.27 hours. BOD and suspended solids removal in the primary clarifiers averages 30 and 55% respectively, according to the plant operators. The primary clarifier effluent is lifted by three screw

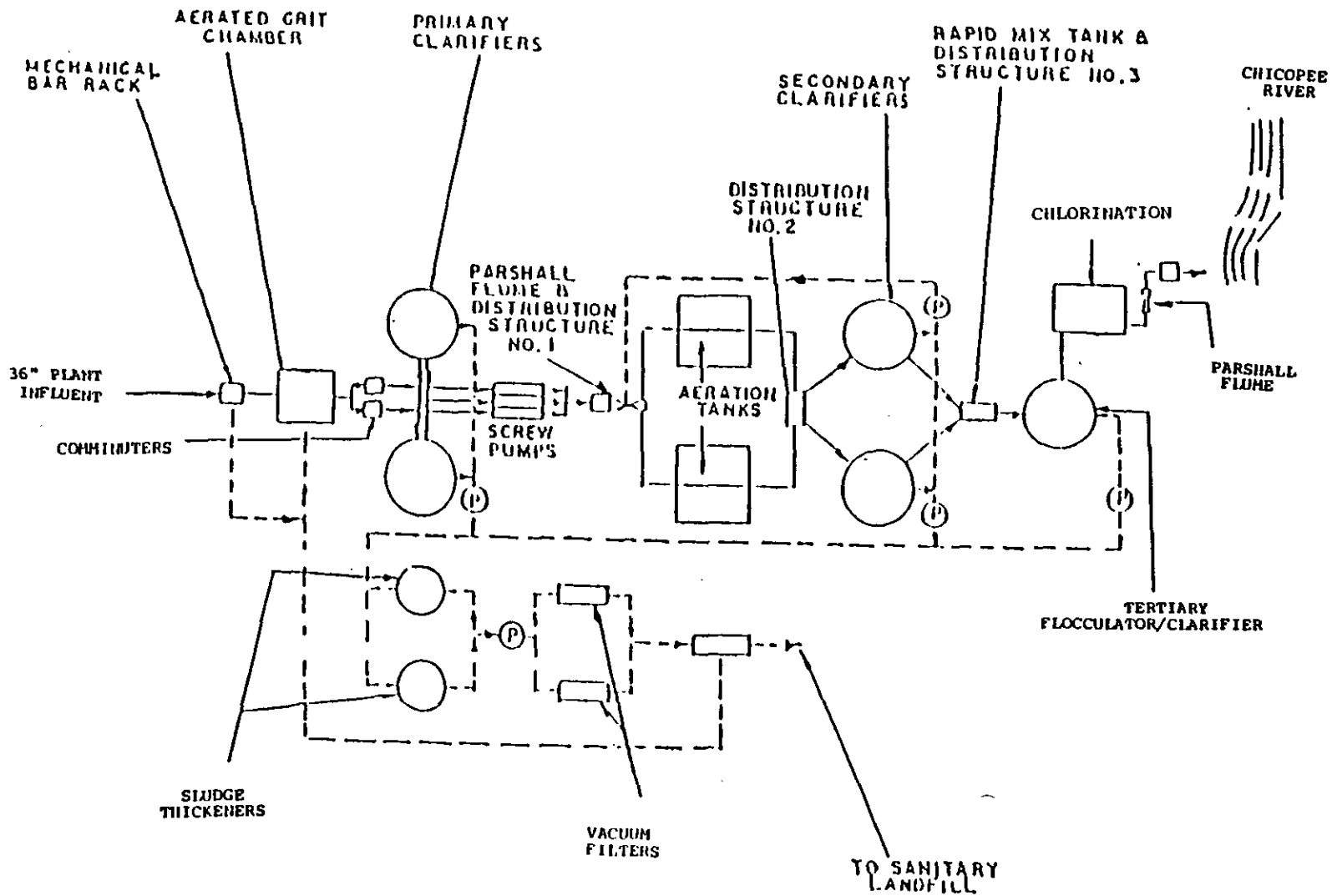


Figure 6. Palmer Wastewater Treatment Facility Schematic Diagram.

pumps to a parshall flume and distribution box before it enters the aeration basins.

There are four aeration basins at the facility which are divided into two trains of two basins each. Eighty submerged static helix bubble air diffusers provide oxygen to each basin. By design, the average hydraulic detention time is 4.4 hours, but the system is currently operated at a detention time of approximately 8.4 hours. The plant was designed for an aeration basin influent BOD load of 140 mg/l although the actual loading is closer to 180 mg/l. Dissolved oxygen levels are maintained slightly above the 2 mg/l design minimum. The aeration basins are operated with a mixed liquor suspended solids concentration of approximately 1600 mg/l in the summer and 2400 mg/l in the winter.

The Palmer facility has two secondary clarifiers 26 meters (85 ft) in diameter and 3 meters (10 ft) deep. At design flow, the overflow rate is, therefore  $20.1 \text{ m}^3/\text{m}^2/\text{d}$  ( $493 \text{ gpd}/\text{ft}^2$ ). The recycle ratio of activated sludge to the front of the aeration basins is typically 20%. Despite the 3 meter (10 ft) clarifier side wall height, the sludge blanket is normally kept between .3 and .6 meters (1 and 2 ft). The underflow solids concentration is typically 1% by weight.

Effluent from the secondary clarifiers is sent to a rapid mix tank and distribution box where alum is added for phosphorus removal. The resulting chemical sludge is then settled in a separate tertiary clarifier having the same dimensions as the

secondary clarifiers, and thus, an overflow rate of  $40.2 \text{ m}^3/\text{m}^2/\text{d}$  ( $986 \text{ gpd}/\text{ft}^2$ ). This section of the plant was originally designed to remove colloidal materials from forest product industry wastes which are found in the plant influent. The tertiary clarifier effluent is chlorinated before passing through a parshall flume for flow measurement and discharge to the Chicopee River.

Sludges from the primary, secondary and tertiary clarifiers are mixed and dosed with  $0.34 \text{ kg}$  ( $3/4 \text{ lb}$ ) of ferric chloride and  $4.5 \text{ kg}$  ( $10 \text{ lb}$ ) of lime for every  $45 \text{ kg}$  ( $100 \text{ lb}$ ) of mixed sludge. The conditioned sludge is gravity thickened before being dewatered by vacuum filtration. According to plant operators, the coil filters work well producing a filter cake which is typically 18 to 20% solids. Unfortunately, the coil filters are energy intensive, and, therefore, expensive to operate. The filter cake is loaded with bar scrapings and grit from preliminary treatment into trucks for final disposal at the town landfill.

The Palmer facility has a seasonal phosphorus permit limitation, but it is not required to convert ammonia or remove nitrogen. The summer season, when phosphorus removal is required, is defined as May 1 through November 1. Palmer effluent limitations and permit requirements are presented in Table 10. The numbers shown in the table are monthly average effluent limits and yearly average influent values.

The plant runs well and is said to be quite flexible in operation. The contribution of the forest products industries in



the area do not seem to have a negative impact on plant performance, and colloidal material in the plant effluent has not been a major problem. These industries often contribute 40% of the plant's organic loading. In contrast to the Pittsfield plant, the alkalinity is low. Alkalinity averages approximately 100 mg/l and varies widely depending on the mix of industrial and domestic wastewater entering the plant. Historically this lack of alkalinity has not caused operational difficulties.

TABLE 10: Palmer Permit and Influent Values

Parameter	Yearly Average Influent	Permit Limit (ave monthly)
BOD <sub>5</sub>	250 mg/l	30 mg/l
Suspended Solids	200 mg/l	30 mg/l
Phosphorus	6.0 mg/l	1.0 mg/l
Nitrogen	12.0 mg/l	none
Flow	0.15 m <sup>3</sup> /s (3.5 mgd)	0.25 m <sup>3</sup> /s (5.6 mgd)

Flows have been known to increase dramatically over short time periods, but, surprisingly, washout has generally not occurred. In one instance, plant operators were able to treat flows in excess of

0.53 m<sup>3</sup>/s (12 mgd) for a one week period and still produce an effluent near permit values. During normal operation, effluent BOD and suspended solids concentration are approximately 10 mg/l, though the suspended solids levels are slightly higher in the winter. Effluent phosphorus levels average between 0.5 and 0.75 mg/l.

The Palmer facility appears to be a good candidate for either the Phostrip or A/O systems. As there is no nitrogen limitation in the plant's permit, the Bardenpho process would not be likely to provide a cost effective alternative. There are a number of reasons why biological phosphorus systems appear to be a good match with this particular plant.

The flexibility of the plant, particularly in respect to dissolved oxygen control in the aeration basin, is an important consideration. The plant operators are very confident of their ability to maintain dissolved oxygen levels at a given concentration over a wide range of influent conditions and solids retention times.

The fact that the plant is now operating with a relatively shallow sludge blanket is another advantage. It appears likely that sludge in the secondary clarifier could be maintained in an aerobic condition without process modification. Similarly, vacuum filtration should not cause a phosphorus release, though some process modification may be necessary in the gravity thickening step.

If the Phostrip process were to be used, the tertiary clarifier or one of the gravity thickeners might be retrofitted for use as a stripper tank. As the plant is already using lime, chemical storage

and handling facilities would be expected to require minimal modification.

Influent phosphorus to BOD ratios also do not seem to present any problems. With average influent phosphorus concentrations in the 5 to 7 mg/l range, excess uptake should be capable of removing sufficient phosphorus to meet the 1 mg/l effluent limit with either process.

Finally, the plant operators expressed a feeling that biological phosphorus removal may be a good alternative for their plant. This operator cooperation may prove to be invaluable in a retrofit of the plant.

4.3 North Attleboro Wastewater Treatment Facility. The first North Attleborough wastewater treatment facility began operation nearly seventy-five years ago and the plant has been upgraded and redesigned in 1945, 1958 and 1977. The present facility was designed by Whitman and Howard Inc. and began operation in March 1980. The facility was designed to treat an average flow of 0.20 m<sup>3</sup>/s (4.6 mgd) and meet the treatment needs of the towns of North Attleborough and Plainville through the year 2000. Flow currently averages around 0.1 m<sup>3</sup>/s (2.3 mgd) in the summer with winter and spring flows being slightly higher. The facility is depicted schematically in Figure 7. Effluent is discharged to the Ten Mile River which is a small stream that serves as a reserve public drinking water source in Rhode Island.

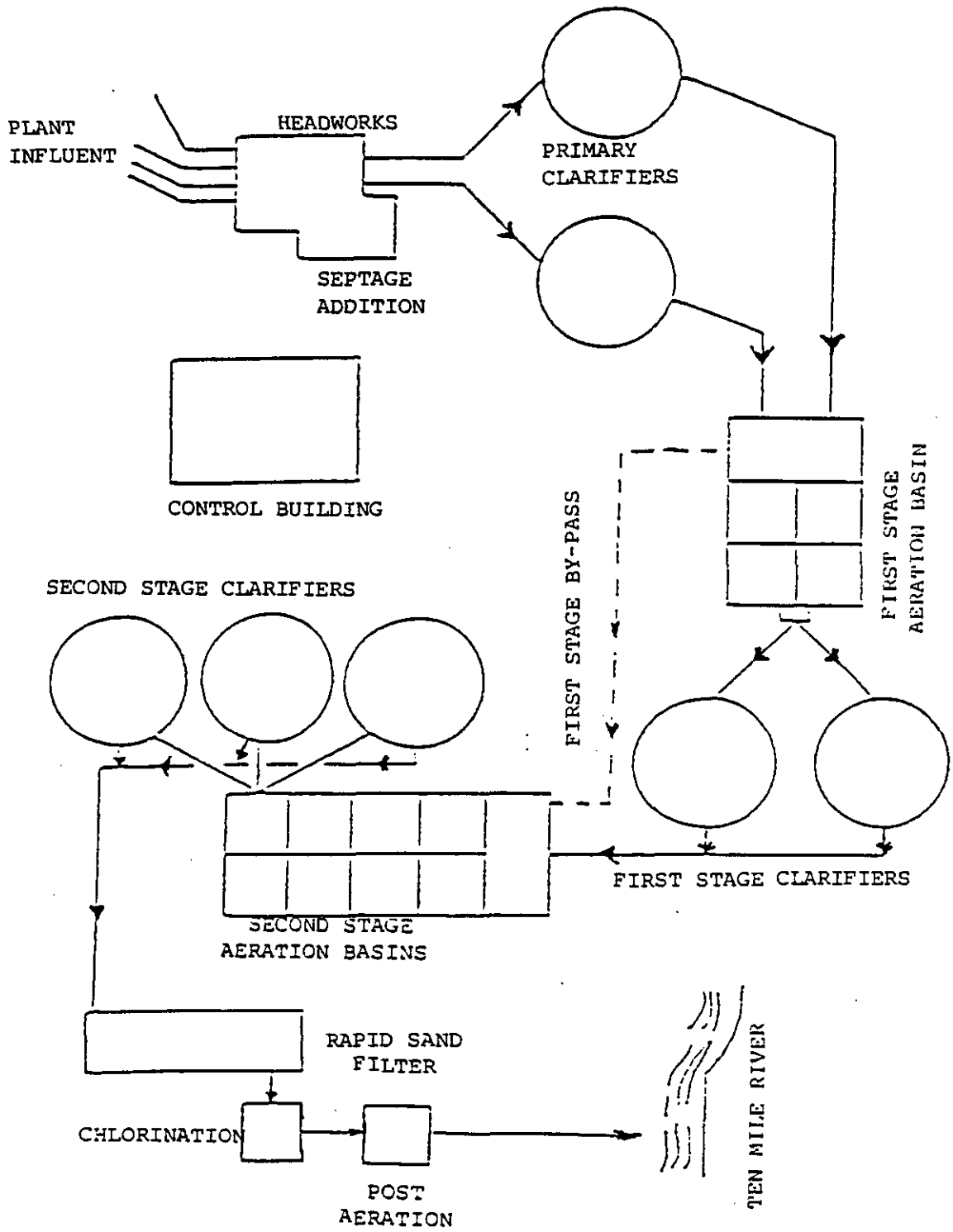


Figure 7. North Attleboro Wastewater Treatment Facility Schematic Diagram.

The plant headworks consist of a mechanically cleaned bar screen, two aerated grit chambers, a parshall flume, two flash mix tanks and two flocculation tanks. The flash mix tanks were designed for alum addition to provide phosphorus removal in the primary clarifier, but this method is not currently employed. Ferrous sulfate is currently being added to the grit chamber at a rate of 380 to 475 l/d (100 to 125 gpd) to improve settling and enhance phosphorus removal. The flash mixers and flocculation tanks are not being used. In addition, two septage holding tanks and two septage pumps were provided to equalize this contribution to the plant influent.

Two circular tanks 24.5 meters (80 ft) in diameter and 3.65 meters (12 ft) deep serve as primary clarifiers though only one is normally used. At design flow using both tanks, the overflow rate is  $21.4 \text{ m}^3/\text{m}^2/\text{d}$  (525 gpd/ft<sup>2</sup>) and the detention time is 4.7 hours. Waste activated sludge is added to the primary clarifiers for thickening. BOD and suspended solids removal in the primary clarifier is generally 50 to 60%. The high rate of removal is probably due partially to the ferrous sulfate addition. The clarifiers are set into the ground so that lift pumps are required to send the clarifier effluent to the remainder of the plant.

The North Attleborough facility was designed as a two stage nitrification system with the ability to bypass the first stage activated sludge process in the winter when ammonia conversion is not required. In practice, the first stage is not used and the flow

is directed to the second stage aeration basin even during periods when nitrification is needed. This unused portion includes four first stage aeration tanks with a total volume of 1225 cubic meters (43,200 ft<sup>3</sup>), two first stage clarifiers with a total volume of 3000 cubic meters (106,080 ft<sup>3</sup>) and a pump station.

The second stage aeration basin is made up of eight tanks 12.2 by 12.2 by 3.67 meters (40 by 40 by 12 ft). According to design, the detention time should be 3.7 hours in a step aeration mode when nitrification is not required and 6.0 hours when the two stage system is used. In practice, a plug flow mode is used with a detention time between 6 and 7 hours in the summer and a contact stabilization mode is used in the winter with a detention time of 2.5 to 3.0 hours. An automatic lime feed system was originally included to adjust pH prior to the second stage aeration basin. This system was found to be oversized and difficult to operate and is not used at present. A small dose of sodium bicarbonate is presently being added for this purpose. Dissolved oxygen levels in the aeration basin are normally maintained in the range of 0.5 to 1.0 mg/l in order to minimize power consumption. The mixed liquor suspended solids concentration is typically 4500 mg/l.

The plant has three circular tanks 24.5 meters (80 ft) in diameter and 3.67 meters (12 ft) deep for secondary clarification. Only two of these units are currently in use. At design flow using all three tanks, the overflow rate is 14.3 m<sup>3</sup>/m<sup>2</sup>/d (350 gpd/ft<sup>2</sup>) and the average detention time is 7 hours. The plant

retains a high solids loading and sludge blanket depths in the secondary clarifier are often 2.1 to 2.7 meters (7 to 9 ft). Sixty to seventy percent of the activated sludge settled in the clarifier is returned to the head of the second stage aeration basin.

The clarifier effluent is then dosed with alum to remove any remaining phosphorus before being sent to a pair of rapid sand filters. The alum used and sludge produced is minimal, probably due to significant phosphorus removal in the primary clarifier. The alum dose is generally 275 to 300 liters per day (70 gallons per day). The resulting sand filter sludge/backwash is sent to the plant headworks where it is settled in the grit chamber.

The plant effluent is then chlorinated in two contact chambers and reaerated in a separate post aeration basin before discharge to the Ten Mile River.

Waste activated sludge from the second stage clarifiers averages only 0.1 to 0.2% solids. This sludge is thickened in the primary clarifier although 2 air flotation thickeners were included in the plant for this purpose. The flotation thickeners were found to require close attention and were too labor intensive to be practical for the plant personnel available.

Sludge from the primary clarifier, including the waste activated sludge and chemical sludge generated from the ferrous sulfate addition is sent to two sludge holding tanks 6.9 by 8.8 by 4.0 meters (22.6 by 22.8 by 13 feet). The tanks are equipped with a lime feed system which is not being used.

Sludge from the holding tanks is mixed with polymer as a dewatering aid at the rate of 7.5 kg per metric ton of dry sludge (15 lb/ton). This translates into a polymer dosage of approximately 91 kg (200 lb) per week or roughly two-thirds of the 137 kg (301 lb) per week of polymer specified by design. Dewatering is accomplished with two centrifuges. The centrifuges increase the solids concentration from slightly under 1% to approximately 20%.

Sludge from the centrifuges and grit and screenings from preliminary treatment are disposed of in the town landfill.

The North Attleborough facility is currently required to both remove phosphorus and convert ammonia nitrogen between June 1 and September 30. BOD and suspended solids limits are also lower for this part of the year. Permit limitations and typical influent values are presented in Table 11.

Phosphorus removal is currently accomplished by chemical addition at two distinct points in the plant. Alum is added specifically to remove phosphorus prior to rapid sand filtration while ferrous sulfate is added to the grit chamber principally to improve settling. In the past, the plant was operated without the ferrous sulfate addition. Based on this experience, plant operators feel that phosphorus removal adds only a small amount of sludge, perhaps 227 to 273 kg (500 to 600 lb), dry weight, per day. This represents 13 to 15% of the 1727 kg (3800 lb) of dry solids produced daily.



TABLE 11: North Attleborough Permit and Influent Values

Parameter	Average Influent (mg/l)	Permit Limit (ave month - mg/l)
BOD <sub>5</sub> (summer)	210	5.0
BOD <sub>5</sub> (winter)	160	15.0
Susp. Solids (sum.)	250	7.0
Susp. Solids (win.)	200	30.0
Phosphorus (summer)	14.0	1.0
Ammonia - N (summer)	10.0	1.0
Flow m <sup>3</sup> /s	0.105 m <sup>3</sup> /s  (2.4 mgd)	0.25  (5.6 mgd)

The plant routinely produces a high quality effluent with phosphorus, suspended solids and BOD concentrations less than half permit limits. Ammonia levels occasionally exceed permit levels, particularly when the influent alkalinity is low. Influent alkalinity concentrations are generally between 80 and 120 mg/l. The addition of sodium bicarbonate to the second stage aeration basin has lead to more stable nitrification and, therefore, better ammonia conversion.

The area has a large number of metal plating industries and a study of pretreatment needs was conducted as a part of the last design contract. Most of the plating industries are currently practicing pretreatment so that toxic metals do not generally pose a problem for the plant's biological treatment units. An area of future concern is the probable imposition of a 20 microgram per liter copper limitation on the plant's effluent. The influent normally contains approximately 100 micrograms per liter.

Inflow and infiltration to the system can be a problem for the facility as plant flows above 5 mgd are common in periods of wet weather. Flow is usually diminished after a week or two, and although the system experiences occasional upset, operation rapidly returns to normal.

Another problem for plant operators is sludge disposal. The plant is only allowed to dispose of two truckloads of sludge per day at the town landfill. This fact has compelled operators to hold large volumes of sludge on site in the sludge holding tanks and the second stage clarifiers.

The plant is very flexible in operation, as evidenced by the many process modifications currently being used. Many of these changes have been necessary to reduce labor and energy costs. In addition, many process modifications have been tried for short periods of time in an attempt to further optimize operation. The plant operators feel that they have good control over dissolved oxygen levels throughout the process. The fiberglass aerators can

be adjusted by raising or lowering them into the wastewater, changing speed, and turning off one or more aerators for varying time periods as they are equipped with timers.

The plant has experimented with the A/O process. In the summer of 1980, plant operators ran the plant according to the A/O process focusing primarily on achieving the oxygen profiles necessary for phosphorus release and uptake. The results were encouraging, but not conclusive and the operators intend to experiment further with the A/O process. During the former experimental period, phosphorus removal was largely accomplished by alum addition prior to rapid sand filtration. Based on their experience with the process, the plant operators feel that the A/O process may represent a cost effective alternative to chemical precipitation at North Attleborough.

The Phostrip and Bardenpho systems may also be applicable to this facility although the Bardenpho system would likely be less cost effective due to its long detention times and aeration requirements. Denitrification is not required at this facility. The plant at present appears to have adequate tankage and flexibility to operate any of the three processes.

One problem which needs to be solved before any biological phosphorus removal process can function well is the solids loading at the facility. Biological phosphorus removal processes themselves may be helpful in reducing this problem as little or no chemical sludge would be produced. The use of the unused second stage clarifier might also be helpful in reducing sludge blanket depths

and maintaining an aerobic sludge blanket.

As previously mentioned, dissolved oxygen control should not present a problem for this plant and the anaerobic/aerobic profiles typical of the A/O process have been achieved in the past.

Sludge thickening and dewatering facilities in place at the plant would be compatible with biological phosphorus removal systems although process modification may be required. The practice of thickening waste activated sludge in the primary clarifier in particular would likely be problematic with the A/O system. In addition, operators would have to be careful to prevent phosphorus release in the sludge holding tanks with the A/O system.

The sludge holding tanks may prove to be ideal candidates for retrofit as stripper tanks if the Phostrip process is to be used. A lime feed system is in place and the resulting calcium phosphate could be precipitated in either the primary clarifier or the unused first stage clarifiers.

With influent phosphorus concentrations in the 15 mg/l range, a small chemical addition may be needed prior to sand filtration in order to meet the 1 mg/l effluent limit, particularly with the A/O system.

Plant operators have demonstrated their ability to make the modifications and adjustments necessary to run a complicated system. This commitment would be a great advantage if a retrofit to any of the biological phosphorus removal systems is to be accomplished.

## C H A P T E R V

## Discussion

5.1 Costs and Economic Considerations. The cost of a treatment system is of primary importance when deciding between systems capable of producing approximately equal effluents, and the decision to use a given process is rarely made without first considering the best cost estimates available. Such estimates are often difficult to obtain and are, at best, approximate. Any such estimate should, therefore be used with care and, then, only as a guide or a decision making tool.

A widely recognized itemized cost breakdown is presented in the report Emerging Technology Assessment of Biological Phosphorus Removal prepared by Roy F. Weston Inc. for the Wastewater Research Division of EPA (Weston, 1985). The report considers the costs for phosphorus removal to 1 or 2 mg/l with and without nitrogen conversion and nitrogen removal by the Phostrip, A/O, Bardenpho, and conventional precipitation methods. While the cost breakdowns given are helpful, their use is somewhat limited for the purposes of this study by the fact that only new plants with flows of 0.022, 0.22 and 2.2 m<sup>3</sup>/s were considered. The cost to retrofit an existing facility would be expected to be quite different from the costs to build a new facility. One must consider what existing equipment can be used with and without modification, the cost of necessary modifications, and the cost of needed new equipment. For these

reasons, cost determinations for retrofitting existing plants to biological phosphorus removal are highly site specific and do not directly lend themselves to the type of cost analysis presented in the Emerging Technology Report.

An example, taken from the Emerging Technology Assessment of Biological Phosphorus Removal report (Weston, 1985), may be used to demonstrate how existing conditions at a given facility may drastically affect costs (see Table 12). The table and this discussion are, of course, valid only for the purpose of comparison.

In applying the costs shown in the table to the North Attleborough facility discussed in Chapter 4, it should be noted that the facility has equipment in place which would make a retrofit much less expensive. If the Phostrip system were used to meet the plant's 1 mg/l effluent limit, and one of the existing treatment units were modified for use as a stripper tank, the \$1,096,000 cost would likely be closer to one-quarter this value, or perhaps \$250,000. Therefore, the total cost (including the other equipment which is shown to maintain the integrity of the comparison) would be \$17,304,000. This is less than cost for conventional precipitation. Similarly, if the A/O system were used, an effluent filter would not be required as the plant now operates a rapid sand filter for effluent polishing. The resulting savings of \$1,300,000 would make the total A/O system cost \$18,289,000, a value which is much more competitive with the conventional precipitation system cost of \$17,757,000. While it is certainly true that the cases given above are not truly representative of the situation at North

TABLE 12: Cost Comparison 5.0 mgd Facility<sup>1</sup>  
Case 1. Phosphorus Removal (Effluent TP=1 mg/l)  
ENR Index = 3875

Process Unit	One - stage activated sludge with alum (baseline)	Phostrip	A/O
Low Lift Pumping	\$ 705,000	\$ 705,000	\$ 705,000
Prelim. Treatment	196,000	196,000	196,000
Primary Treatment	438,000	438,000	438,000
Aeration/Clarification	2,035,000	2,035,000	2,129,000
Phostrip (with lime)	---	1,096,000	---
Chemical Addition	78,000	---	12,000
Nitrification/Clarif.	---	---	---
Denitr./ Clarification	---	---	---
Chlorination	191,000	191,000	191,000
Effluent Filtration	---	---	1,300,000
Gravity Outfall	310,000	310,000	310,000
Misc. Structures	250,000	250,000	250,000
Thickening (DAF)	204,000	188,000	196,000
Digestion (Aerobic)	470,000	470,000	500,000
Dewatering (Vac Filt)	658,000	595,000	595,000
Sludge Haul/Landfill	251,000	246,000	247,000
Sub - Total	\$ 5,786,000	\$ 6,720,000	\$ 7,069,000
Noncomponent Cost <sup>2</sup>	1,620,000	1,882,000	1,979,000
Engineering and Constr. Supervision	1,111,000	1,290,000	1,357,000
Contingency	1,111,000	1,290,000	1,357,000
TOTAL CAPITAL COST	\$ 9,628,000	\$ 11,182,000	\$ 11,763,000
ANNUAL O&M COSTS	\$ 805,000	\$ 690,000	\$ 775,000
TOTAL PRES WORTH COST <sup>3</sup>	\$ 17,757,000	\$ 18,150,000	\$ 19,589,000

1. This table is taken from TABLE B-2 of the report Emerging Technology Assessment of Biological Phosphorus Removal (Weston, 1984)

2. Noncomponent Costs include piping, electrical, instrumentation, and site preparation.

3. Present Worth computed assuming a 20 year life at a 7-5/8% interest rate (PWF = 10.0983).

Attleborough, both because many unnecessary costs are included and the design flow is actually  $0.20 \text{ m}^3/\text{s}$  rather than  $0.22 \text{ m}^3/\text{s}$ , the point remains that retrofit costs are extremely site dependent.

The major capital costs associated with retrofitting a facility to the Phostrip system are the stripper tank, the lime storage and feed system, additional pumping (as required), additional instrumentation and control equipment (if needed), and licensing fees. Of these the stripper tank generally represents the largest portion of the total capital cost unless an existing equipment unit can be modified for this purpose.

In order to retrofit a facility to the A/O system, the major capital costs involved are additional aeration equipment (if required), partitions to divide the aeration basin into sections, covers for the anaerobic zone, mixers for the anaerobic zone, instrumentation and control equipment (if needed), and licensing fees.

The major capital costs involved in retrofitting a facility to chemical precipitation are a chemical storage building, a chemical mixing and feed system, a flocculation tank (if required), a sedimentation basin (for a tertiary system), and additional instrumentation and control equipment. Many of these costs would also be dependent on existing equipment as well as the point of addition and the chemical chosen.

Approximate "level one" cost estimates for retrofitting the North Attleborough and Palmer facilities were prepared by Biospherics and Air Products based on the information contained in



sections 4.2 and 4.3 of this report. While these estimates are approximate, they are useful as they give the best indication of the magnitude of the costs to be expected in retrofitting typical Massachusetts municipal wastewater treatment plants to biological phosphorus removal.

For the Palmer facility, Air Products has estimated the necessary capital costs for conversion to the A/O system to be \$508,000. Of this, nearly 90% (\$448,000) is the licensing or "technology" fee. Staging walls to divide the aeration basin into aerobic and anaerobic sections were estimated to cost \$20,000 while eight submerged mixers for the anaerobic sections were estimated to cost \$40,000. Costs for removal of the present aerators and covers for the anaerobic sections were not included in the estimate prepared by Air Products.

In order to retrofit the North Attleborough facility to biological phosphorus removal and nitrification using the A/O system, Air Products has estimated that staging walls would cost \$10,000, mixers \$20,000, and the "technology" fee would be \$448,000 for a total retrofit cost, excluding removal of existing equipment and anaerobic section covers, of \$478,000. In addition, Air Products expects that while a small supplemental alum or ferrous sulfate dose would be required in order to meet North Attleborough's permit limits, a better settling sludge would be expected with the A/O system. The cost estimate prepared by Air Products is included as Appendix 1.

Biospherics listed seven cost elements to be considered in retrofitting either plant:

1. Conversion of existing tankage to strippers,
2. lime storage and feed system modifications,
3. piping modifications to incorporate stripper flows,
4. design engineering, .
5. hardware provided by Biospherics,
6. technical assistance provided by Biospherics,
7. license fee.

Costs for items 5 through 7 for either plant have been estimated by Biospherics to be between \$350,000 and \$700,000. In both cases, the costs of retrofitting existing clarifiers to stripper tanks would be minimal as sludge withdrawal equipment and sludge pumps are expected to be useable with little modification. In addition, both plants have existing lime feed systems which may be useable with minor modification, thus further reducing costs. License fees would be \$184,000 for North Attleborough and \$224,000 for Palmer, based upon a rate of \$40,000 per mgd of plant capacity.

Based upon annual cost savings experienced at Reno/Sparks, Nevada of \$15,000 to 20,000 per year per mgd (Peirano, 1977), Biospherics has estimated a 4 to 10 year payback at North Attleborough and a 3 to 8 year payback at Palmer. The cost estimate prepared by Biospherics appears in Appendix 2.

In order to develop a more meaningful "design life" type of comparison, an attempt has been made to combine capital costs with operation and maintenance costs for both the Palmer and North

Attleborough facilities. In each case, capital costs given by Air Products and Biospherics have been used at "face value". This is to say it is assumed that each plant can be successfully retrofitted for the dollar amounts estimated. An average figure of \$500,000.00 has been used for retrofitting both plants to the Phostrip system.

In addition, the following assumptions were made:

- (1) A 20 year planning horizon was used,
- (2) the discount rate used was 8% ( $PWF = 9.81815$ ),
- (3) operation and maintenance costs exclusive of chemical inputs were considered equal between the three alternatives,
- (4) alum costs are based on liquid alum containing 48.5% aluminum sulfate. Alum cost used was \$12.00 per 100 pounds of product delivered. Liquid alum density used was 11.2 lb/gallon,
- (5) ferrous sulfate solution is assumed to be 55% ferrous sulfate and having a density of 65 lb/ft<sup>3</sup>. Costs are based on purchasing dry ferrous sulfate (75%  $FeSO_4 \cdot 7H_2O$ ) at \$25.00 per 100 pounds,
- (6) lime costs were based on a price of \$6.00 per 100 pounds, bulk delivered,
- (7) lime usage for the Phostrip processes was calculated using a dose of 200 mg/l (due to low alkalinity) and assuming the elutriate flow is 13% of the plant's forward flow,
- (8) a small dose of alum capable of precipitating 3 mg/l of phosphorus has been assumed for the A/O alternative at North Attleborough due to the high influent phosphorus concentration.

The comparison thus developed is shown in Table 13. From the table it can be seen that the A/O system shows the lowest overall cost for Palmer while Phostrip appears to be the least cost alternative for North Attleborough. More importantly, the table shows that both biological systems can be competitive with conventional precipitation. It cannot be overemphasized that the costs shown in the table have been developed for comparison use in this study only. An other use would be, at best, imprudent. The calculations used to generate the figures shown in Table 13 are given in Appendix 3. It is worthy of note that the conventional precipitation alternative shown for North Attleborough is actually a specialized treatment system developed by the operators while the conventional alternative shown for Palmer is based upon design dosage found in the literature, thus comparisons between plants in particular should be avoided.

5.2 Operation and Performance. Some elements of the engineering community have remained skeptical of the ability of biological phosphorus removal systems to consistently meet effluent limitations. To a large degree, this is attributable to the fact that biological phosphorus removal is a new technology and effluent phosphorus limits have been changed often in recent years. In addition, a number of problems have been experienced in all three types of plants, often causing unstable performance or delaying the start up of a system. The problems encountered do not appear to be

TABLE 13: Preliminary Cost Comparison  
for Alternate Systems at North Attleborough and Palmer.

	Capital Cost (Present Worth)	Chemical Cost (Annual)	Chemical Cost (Present Worth)	Total Cost (Pr. Worth)
PALMER:				
A/O	\$508,000	-	-	\$508,000
Phostrip	500,000	\$7,000	\$69,000	569,000
Conv.	-	106,000	1,040,000	1,040,000
NORTH ATTLEBOROUGH:				
A/O	478,000	30,200	296,000	774,000
Phostrip	500,000	3,800	37,000	537,000
Conv. (Present Method)		55,000	540,000	540,000

DISCLAIMER: The costs shown have been developed for the purposes of comparison only.

related to deficiencies in the processes themselves, however, but are generally related to mechanical design and equipment selection (Weston, 1984).

The problems encountered at the Amherst, New York facility illustrate some of the problems to be avoided when designing a biological phosphorus removal system. The Phostrip process was chosen for the facility when the plant was 80% constructed; a fact which accounts for many of the operational problems and mechanical difficulties experienced at the plant. Mechanical difficulties developed with the variable speed drives for the stripper and

reactor/clarifier tanks, the agitator shaft on the lime feed line, the fiberglass make up tank for the lime feed slurry, and the automatic pH control - lime feed instrumentation system (Northrup and Smith, 1983). In addition, long piping runs made sludge recycle pumping both expensive and difficult when winter conditions make freezing a problem. Despite these problems, some of which caused the lime feed system to be inoperative between December 1981 and March 1982, plant operators found that the Phostrip system was able to significantly reduce effluent phosphorus concentrations even without chemical precipitation of the stripper supernatant (Northrup and Smith, 1983). The problem which ultimately led to the discontinuation of Phostrip use at Amherst was the formation of calcium carbonate on critical exposed surfaces. This was caused by the build up of carbon dioxide from microbial respiration in the covered pure oxygen aeration basin.

Despite the problems experienced with the start up of some facilities, the Phostrip system has demonstrated its ability to produce effluent concentrations below 1 mg/l. Monthly influent and effluent values for the Little Patuxent Wastewater Treatment Plant in Savage, Maryland are shown in Table 14.

Problems have also been experienced at the Palmetto, Florida Wastewater Treatment Plant using the Bardenpho system. Clogging of the primary clarifier underflow and dissolved oxygen control problems led to poor phosphorus removal during the plant's initial

TABLE 14: Phostrip Performance, Little Patuxent  
Wastewater Treatment Plant, Savage, Maryland

Month, Year		Plant Influent Tot. Phos. (mg/l)	Sec. Effluent Tot. Phos. (mg/l)	Final Effluent Tot. Phos. (mg/l)
Nov, 1984	Max	10.9	3.2	1.1
	Min	7.1	0.1	0.1
	Ave	8.9	0.9	0.3
Dec, 1984	Max	14.1	1.1	0.3
	Min	7.8	0.2	0.1
	Ave	9.6	0.5	0.2
Jan, 1985	Max	13.4	1.3	1.1
	Min	6.9	0.2	0.2
	Ave	9.7	0.6	0.4
Feb, 1985	Max	22.5	3.6	2.2
	Min	5.0	0.3	0.2
	Ave	8.9	1.3	0.8
March, 1985	Max	13.0	2.3	1.2
	Min	7.9	0.5	0.2
	Ave	9.5	1.0	0.5
Five Month Average			0.9	0.4

Bardenpho operation. Though an effluent phosphorus concentration capable of meeting the plant's 1 mg/l permit limit was not found to be sustainable, the Bardenpho system did demonstrate the ability to remove 98% of the influent nitrogen while removing 65% of the influent phosphorus, 98.9% of the influent BOD, and 98.6% of the influent suspended solids. Performance of the Palmetto facility for the period June 1983 to November 1983 is shown in Table 15 (Eimco Process Equipment Company, 1984).

TABLE 15: Bardenpho Performance  
Palmetto Wastewater Treatment Plant  
Palmetto, Florida

Month, Year	Influent Tot. Phos. (mg/l)	Influent Tot. Nitr. (mg/l)	Effluent Tot. Phos. (mg/l)	Effluent Tot. Phos. (mg/l)
June, 1983 Ave.	6.0 <sup>1</sup>	35.0 <sup>2</sup>	1.7	2.4
July, 1983 Ave.	6.0	35.0	1.2 <sup>2</sup>	2.0
August, 1983 Ave.	6.0	35.0	0.6 <sup>3</sup>	2.5
September, 1983 Ave.	6.0	35.0	0.5 <sup>3</sup>	2.5
October, 1983 Ave.	6.0	35.0	1.0 <sup>2</sup>	2.4
November, 1983 Ave.	6.0	35.0	0.4 <sup>3</sup>	2.8

Average influent BOD and Suspended Solids were 110 mg/l and 108 mg/l, respectively, for the six month period.  
Average effluent BOD and Suspended Solids were 1.7 mg/l and 1.4 mg/l, respectively, for the six month period.

Key:

1. Figures for influent total phosphorus and nitrogen values shown are six month averages.
2. Chemical was not added for enhanced phosphorus removal.
3. A "minimal" alum dose was added prior to final clarification in order to enhance phosphorus removal.



The start up of the pilot scale A/O system at the Patapsco Wastewater Treatment Plant was hindered by similar mechanical difficulties as those experienced in the Phostrip and Bardenpho systems described above. One such problem was caused by high liquid levels in the aeration units which went undetected for a four week period due to the lack of adequate liquid level and dissolved oxygen control instrumentation. A plug of grease and foam was found to have blocked the clarifier centerwell causing the surface aerators to become partially submerged and, thus, ineffective (Deakyne et al., 1983). On other occasions, the dissolved oxygen concentration dropped to "very low levels", so that an additional oxygen supply was added. Another problem with the Patapsco pilot plant was an inability to vary the return sludge flow adequately. This oversight made it impossible to maintain the desired MLSS concentration. In spite of the problems listed above, the pilot plant was able to achieve stable operation and was able to produce an average effluent phosphorus concentration of 1.4 mg/l for the December 8, 1982 to January 11, 1983 period (Deakyne et al., 1983). This is below the plant's present 1.5 mg/l permit limit. These results were sufficient to convince the city of Baltimore to retrofit the entire 3.0 m<sup>3</sup>/s (70 mgd) facility to the A/O system.

The city of Largo, Florida is currently using the A/O process to meet its nutrient removal needs. The plant is operated in a denitrifying mode, and the plant operators have been pleased with the A/O system performance. It is worthy of note, however, that while chemicals have not been necessary for phosphorus removal, the

plant's permit limit has been raised to 4 mg/l on a yearly average basis. In the month of March, 1984, the plant effluent averaged 1 mg/l ammonia nitrogen, less than 10 mg/l total nitrogen and 3.1 mg/l total phosphorus. Table 16 presents monthly average effluent concentrations for the plant when it was operated in nitrifying and non-nitrifying modes.

TABLE 16: A/O Performance  
Largo Wastewater Treatment Plant  
Largo, Florida

Month, Year	Effluent Total Phos. (mg/l)	Effluent Amonia Nitrogen (mg/l)	Effluent BOD (mg/l)
February, 1981	1.51	9.84 <sup>1</sup>	5.5
March, 1981	1.27	10.93 <sup>1</sup>	6.7
April, 1981	1.17	10.52 <sup>1</sup>	7.9
May, 1981	1.43	12.05 <sup>1</sup>	4.2
June, 1981	1.37	9.89 <sup>1</sup>	8.0
July, 1981	2.09	0.32	4.1
August, 1981	2.06	0.23	4.4
September, 1981	1.41	0.71	---
October, 1981	2.16	1.72	6.7
November, 1981	1.33	2.53	4.2
December, 1981	1.51	1.65	6.2
January, 1982	1.73	0.24	5.1
February, 1982	1.89	1.52	6.5

All values shown are monthly averages. "Typical" influent total phosphorus and total kjeldahl nitrogen were 8.9 mg/l and 29.4 mg/l, respectively. "Typical" influent BOD was 145 mg/l.

Key:

1 System was operating in non nitrifying mode.

## CHAPTER VI

## Conclusions and Recommendations

A careful review of biological phosphorus removal systems leaves one with several impressions and some conclusions. Even though much is still to be learned about biological phosphorus removal and the mechanisms involved, it is clear that biological phosphorus removal is a workable alternative to chemical precipitation for many municipal wastewater treatment plants. This is not to say that biological phosphorus removal systems are likely to replace chemical precipitation. Both technologies will be important in the future, and the best method should be chosen on a site by site basis.

This was found to be the case with the three plants chosen for detailed consideration in this study. The Pittsfield plant was not found to be a good candidate for any of the three biological phosphorus removal systems. The trickling filters and anaerobic digesters employed at this plant would likely make biological phosphorus removal and sludge handling difficult and expensive. The Palmer facility was found to be a good candidate to retrofit using either the Phostrip or A/O systems as plant operation is quite flexible and either system would result in lower costs than those for conventional chemical precipitation. The A/O system produced the lowest overall cost estimate. Similarly, the North Attleborough facility was found to present good opportunities for biological phosphorus removal system retrofit using either the Phostrip or A/O systems. In this case the Phostrip system produced the lowest

overall estimated costs while the A/O system showed the highest overall estimated costs. An additional area of concern with the A/O system at North Attleborough is whether the plant's 1 mg/l phosphorus effluent limit can be met without a supplemental chemical dose. The need for such a chemical dose and its magnitude would have to be determined with pilot scale testing at the plant. The Bardenpho system did not appear to offer an economical alternative at any of the plants studied at this time. This is mainly a function of the long retention times, and consequently, large tankage requirements necessary to achieve the denitrification which is integral to the Bardenpho system.

There are a number of factors which may be helpful in determining which plants would be likely candidates for retrofitting to one of the three proprietary processes currently available in the United States:

1. The plant should be flexible in operation. In particular, operators should be able to easily vary the dissolved oxygen level in the aeration basins and have good control over sludge wasting and sludge recycle,

2. the secondary clarifier should be designed conservatively so that phosphorus rich solids are not released over the weir and low sludge blankets can be maintained. Sludge blanket depths should ideally be less than two feet in order to keep the sludge aerobic and prevent phosphorus release,

3. sludge thickening and dewatering processes should be investigated to insure that anaerobic conditions will not be encountered. For this reason, gravity thickening, aerobic digestion, and anaerobic digestion may cause problems of phosphorus release from the biomass, while dissolved air flotation thickeners, belt filter presses, plate and frame presses, vacuum filters, centrifuges, sand drying beds, and incinerators would not be likely to cause these problems,

4. plant operators should be receptive to the use of a biological phosphorus removal system. Though the systems are not difficult to operate, some modification in operation and procedure will be required, and initial start up can be difficult,

5. if the Phostrip system is used, additional care should be taken when the activated sludge system is supplied with pure oxygen aeration. Though problems were encountered in Amherst, New York, a pure oxygen Phostrip system is working well at Tahoe/Truckee, California.

In considering the use of biological phosphorus removal systems in Massachusetts, a few observations are worthy of consideration. Among them are the following:

1. Phosphorus limits should be made somewhat more standard and changed less often. Plant operators were often unable to recall the exact date that phosphorus removal was required at their facility

and they said that their permits had been changed quite often. Permits could be more standard in terms of whether samples were to be composite or grab samples, and whether the limits were based on maximum discharge or on average concentrations. These changes would certainly simplify the design and operation of both biological and chemical techniques,

2. the seasonal nature of phosphorus limits in the Commonwealth should not present any operational problems although biological systems may not be as cost competitive against chemical precipitation as they might otherwise be. This is due to the fact that the major costs of chemical precipitation (the chemicals themselves) will not be incurred for part of the year while the major costs for biological systems (capital and equipment) will be incurred regardless of use,

3. because denitrification is not required at any of the plants having phosphorus permit limitations in the Commonwealth, the Bardenpho system is not likely to find much application in Massachusetts,

4. in the future, as chemical costs increase and sludge disposal becomes a more pressing and costly problem, the prospects for biological phosphorus removal systems should improve. In addition, if metal leakage from chemical precipitation becomes a concern in acidified waters, biological methods of phosphorus removal will be advantageous.

At this time, it appears that biological phosphorus removal systems can be used to advantage in Massachusetts, and they will likely be increasingly important in the future. Cost savings and additional environmental benefits will result at some plants when care is taken to ensure that plants are retrofitted with adequate flexibility. For the near future, however, chemical precipitation is likely to be the most often used technology for phosphorus removal in municipal wastewater treatment plants within the Commonwealth of Massachusetts.



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A P P E N D I X I

BIOSPHERICS COST ESTIMATES

Design Memorandum:

Retrofit for PhoStrip Biological Phosphorus Removal  
At the North Attleboro, Mass. Wastewater Treatment Facility  
And the Palmer, Mass. Wastewater Treatment Plant

Prepared for Mr. Clayton Richardson

Prepared by  
Biospherics Inc.  
Rockville, Maryland  
14 November 1985

## I. Design Concepts

- A. The overriding consideration in the design of the PhoStrip installations at North Attleboro and Palmer is to develop the simplest PhoStrip design configuration to minimize retrofit costs. Such a configuration will also require minimal disruption of the treatment systems.
- B. For optimum stripper performance and operational flexibility, the stripper and associated piping should be designed to operate with each treatment plant receiving 100% of its design flow, with consideration of and allowance for operation at as low as 50% of plant design flow.
- C. Modifications to the following system components will be required:
  1. Piping modifications must be made to:
    - a. Route a portion of the clarifier return sludge to the stripper at each plant.
    - b. Convey stripper underflow sludge to the aeration basins.
    - c. Recycle stripper underflow sludge to the top of the stripper.
    - d. Convey stripper supernatant overflow to the point of chemical addition, and then to the primary clarifiers.
    - e. Convey lime slurry to the point of chemical addition to the stripper supernatant.
  2. Based on the design criteria and dimensions presented in Section II, one of the first stage clarifiers at North Attleboro and the tertiary clarifier/flocculator at Palmer can serve as stripper tanks for the respective plants. The stripper tanks should be equipped with:
    - a. A center well for sludge discharge to the stripper;
    - b. A scum baffle and an overflow weir with piping connections to convey supernatant overflow from the stripper;
    - c. A sludge rake mechanism;
    - d. Piping and a sludge pump to provide for recycling underflow sludge back to the top of the stripper.
    - e. Sludge blanket level indicators.
    - f. Underflow solids density probes.
  3. Facilities for lime storage, slaking, and slurry feed exist at each treatment plant. Maximum estimated lime requirements for treating stripper supernatant are 210 lbs/day at North Attleboro and 260 lbs/day at Palmer. Slurry feed rates ( 8% lime slurry) are 170 gpd at North Attleboro and 210 gpd at Palmer. (See Section III, item B. 3.)

4. Lime slurry addition to the stripper supernatant can be accomplished using a static in-line mixing device, or, if available, in a flash-mix chamber (such as the rapid mix structure at Palmer) that provides about one minute of detention time.

## II. Stripper Design

### A. Tank Design Factors, North Attleboro:

1. Anaerobic Retention Period ..... 8 hours
2. Influent Wastewater Flow ..... 4.6 mgd
3. Return Sludge as a Percent of  
Primary Effluent Flow ..... 50 percent
4. Percent of Return Sludge Routed to  
Stripper ..... 50 percent
5. Stripper Concentration Factor ..... 1.5
6. Blanket Density Factor ..... 0.8
7. Depth of Stripper Supernatant ..... 4.0 ft
8. Stripper Tank Diameter ..... 80.0 ft
9. Stripper Tank Depth Required for ARP ..... 4.5 ft
10. Stripper Tank Total Liquid Depth ..... 8.5 ft

### B. Tank Design Factors, Palmer:

1. Anaerobic Retention Period ..... 8 hours
2. Influent Wastewater Flow ..... 5.6 mgd
3. Return Sludge as a Percent of  
Primary Effluent Flow ..... 50 percent
4. Percent of Return Sludge Routed to  
Stripper ..... 50 percent
5. Stripper Concentration Factor ..... 1.5
6. Blanket Density Factor ..... 0.8
7. Depth of Stripper Supernatant ..... 4.0 ft
8. Stripper Tank Diameter ..... 85.0 ft
9. Stripper Tank Depth Required for ARP ..... 5.2 ft
10. Stripper Tank Total Liquid Depth ..... 9.2 ft

Stripper dimensions are calculated by determining the required volume of sludge to achieve the design anaerobic retention period (factor 1) for the given tank diameter (factor 8), and adding depth for supernatant (factor 7). The volume required to achieve the desired anaerobic retention period is determined by first calculating the flow of return sludge from the stripper (factor 2 times factor 3, times factor 4, divided by factor 5) and dividing this flow into the desired anaerobic retention period (time divided by flow gives volume); this volume is then divided by factor 6, the blanket density factor, to adjust for the average blanket density as compared to the density of the stripper underflow sludge. The stripper concentration factor (factor 5) is a measure of the stripper underflow sludge concentration as compared to the clarifier return sludge routed to the stripper. A factor of 1.5 means that the stripper underflow sludge is 50% more concentrated than the clarifier return sludge, so the sludge flow rate out of the stripper will be 2/3 the return sludge flow rate into the stripper for a given mass flux through the stripper.



### C. Internal Stripper Mechanisms

The attached drawing will give guidance for design of the stripper tank internal mechanisms. The drawing is provided for reference only, and does not represent recommended dimensions for the stripper tanks at North Attleboro or Palmer. Also, the drawing shows a stripper tank fitted with an elutriation system. An elutriation system is not recommended for the stripper tanks at North Attleboro or Palmer. The reader should disregard the elutriation system features in the attached drawing.

Rotation of the sludge rake should be 2 to 4 revolutions per hour.

### D. Stripper Piping

Sludge piping should be sized for a minimum velocity of 2.0 ft/sec and a maximum velocity of 8.0 ft/sec.

Table 1 presents the pipe size selection procedure. Flows (b) for each stream (a) into or out of the stripper were calculated for operation at 50% and 100% of plant design flows at North Attleboro and Palmer. Pipe velocities for each flow in a range of pipe sizes were then calculated (Table 2). Maximum (c) and minimum (d) pipe diameters were then selected for each flow based on the criteria described above. An allowable design range (e) was then determined from the largest minimum pipe sizes, at 100% of design flow, and the smallest maximum diameters, which were the maximum diameters at 50% of design flow. The recommended design diameters (f) are at roughly the midpoints of the respective design diameter ranges.

### E. Pump Sizes

The pumps for supplying clarifier return sludge to the strippers should be sized to provide variable flow from 0.5 to 1.2 mgd at North Attleboro and 0.7 to 1.4 mgd at Palmer. The pumps for recycling stripper underflow sludge should be selected to provide variable flows from 0.2 to 0.8 mgd at North Attleboro and 0.2 to 1.0 mgd at Palmer.

### III. Cost Estimates

Most of the retrofit costs will depend on specific site conditions. The following discussions describe the cost components involved in retrofitting for PhoStrip and provide the basis for cost estimations. Where possible, cost elements are given.

#### A. Conversion of existing tankage to strippers:

##### 1. Possible use of internal mechanisms.

It is likely that the internal mechanisms in the tanks to be converted to strippers are adequate, particularly the tertiary clarifier at Palmer. Key features of the stripper internal mechanisms are listed in Section I, item 2.

##### 2. Provision for sludge recycle.

The underflow sludge pump discharge for each converted stripper should include piping and valves to allow stripper sludge recycling.

#### B. Modifications to lime storage and feed systems.

##### 1. Slurry piping is required to the point of addition to stripper supernatant.

##### 2. Storage requirements depend on local availability and bulk purchase discounts. Existing storage may be adequate.

##### 3. Feed requirements estimates are presented below. Existing feed systems may be adequate.

CaO Dose:                    300 #CaO/MG  
                                  65 mg/L

Lime Purity:                55%  
Slurry Concentration:      8 #CaO/100 # slurry  
                                  14.5 #lime/100 # slurry  
                                  0.67 #CaO/gal. slurry

	North	
	Attleboro	Palmer
Max. Supernatant Flow:	0.38	0.47 MGD
Max. Slurry Feed Rate:	171	211 gpd
	7.1	8.8 gph
Max. Lime use	207	256 #/day

Assumes specific gravity of slurry = 1.0 for conservative design.

D. Other piping modifications.

See Table 1 for flows and pipe diameters.

1. Clarifier return sludge to stripper.
2. Stripper underflow to aeration basins.
3. Stripper supernatant to lime addition and primary clarifiers.

E. Design engineering.

Design engineering services will be required for converted stripper tank modifications, piping modifications, and lime system modifications (if necessary).

F. Hardware provided by Biospherics:

1. Stripper internal mechanisms (if necessary).
2. Lime feed systems (if necessary).
3. Instrumentation:
  - a. Sludge density meters.
  - b. Stripper sludge blanket level indicators.
  - c. pH feedback control system for lime addition to stripper supernatant.
  - d. Lime feed control system (if necessary).
  - e. Stripper underflow and recycle pumping control and monitoring system.

G. Technical assistance by Biospherics:

1. Preliminary design assistance to design consultant.
2. Mechanical and process review of consultant's final design.
3. Start-up supervision.  
Hardware: manufacturers  
Process: Biospherics
4. Performance test supervision.
5. Operator training.
6. Operation and maintenance manuals.
7. Service contract for one year after performance testing, renewable annually.

## H. License fees.

North Attleboro.

4.6 MGD @ \$40,000 = \$194,000

Palmer.

5.6 MGD @ \$40,000 = \$224,000

The total estimated costs for Biospherics' scope of services (items F, G, and H) are \$350,000 to \$700,000 for each plant, depending on the extent of modifications to tanks converted to strippers, and the modifications to the existing lime systems. These costs are provided for preliminary planning purposes only, and do not represent a firm price proposal by Biospherics.

TABLE 1. Recommended Pipe Diameters for Stripper Streams.

(a) Stream	50% Design Flow			100% Design Flow			(e) Design Range (in.)	(f) Design Dia. (in.)
	(b) Flow (mgd)	(c) Max. Dia. (in.)	(d) Min. Dia. (in.)	(b) Flow (mgd)	(c) Max. Dia. (in.)	(d) Min. Dia. (in.)		
North Attleboro								
Plant Flow	2.30			4.60				
FEED	0.58	8	6	1.15	12	8	8	8
Max.Tot. FEED	0.96	10	6	1.92	15	10	10	10
SRAS	0.38	6	4	0.77	10	6	6	6
Min. Recycle	0.19	4	4	0.38	6	4	4	4
Max. Recycle	0.38	6	4	0.77	10	6	6	6
Total UF	0.77	10	6	1.53	14	8	8 to 10	10
Supernatant	0.19	4	4	0.38	6	4	4	4
Palmer								
Plant Flow	2.80			5.60				
FEED	0.70	10	6	1.40	14	8	8 to 10	10
Max.Tot. FEED	1.17	12	8	2.33	18	10	10 to 12	12
SRAS	0.47	8	6	0.93	10	6	6 to 8	8
Min. Recycle	0.23	4	4	0.47	8	6	*	6
Max. Recycle	0.47	8	6	0.93	10	6	6 to 8	8
Total UF	0.93	10	6	1.87	16	10	10	10
Supernatant	0.23	4	4	0.47	8	6	*	6

Note: Letters above column headings are text references.

Abbreviations:

FEED ..... Clarifier return sludge flow routed to the stripper  
 Max.Tot. FEED ... FEED plus maximum underflow recycle  
 SRAS ..... Stripper sludge underflow returned to aeration  
 Min. Recycle .... 25% of stripper sludge underflow  
 Max. Recycle .... 50% of stripper sludge underflow  
 Total UF ..... Total stripper sludge underflow  
 Supernatant ..... Stripper supernatant flow

TABLE 2. Pipe Velocities for Selected Flows and Pipe Diameters.

Dia. (in.)	Area (in. <sup>2</sup> )	Velocity (ft/sec)															
		Flow (gpd)															
		0.19	0.23	0.38	0.47	0.58	0.70	0.77	0.93	0.96	1.15	1.17	1.40	1.53	1.87	1.92	2.33
4	12.6	3.4	4.1	6.8	8.3	10.2	12.4	13.6	16.5	17.0							
6	28.3	1.5	1.8	3.0	3.7	4.5	5.5	6.0	7.4	7.6	9.1	9.2	11.0	12.1	14.7	15.1	18.4
8	50.3			1.7	2.1	2.5	3.1	3.4	4.1	4.2	5.1	5.2	6.2	6.9	8.3	8.5	10.3
10	78.5				1.3	1.6	2.0	2.2	2.6	2.7	3.3	3.3	4.0	4.4	5.3	5.4	6.6
12	113.1							1.5	1.8	1.9	2.3	2.3	2.9	3.0	3.7	3.8	4.6
14	153.9										1.7	1.7	2.0	2.2	2.7	2.8	3.4
16	201.1													1.7	2.1	2.1	2.6
18	254.5													1.3	1.6	1.7	2.0

A P P E N D I X 2

AIR PRODUCTS COST ESTIMATES.

Air Products and Chemicals, Inc.  
 Box 536  
 Allentown, PA 18105  
 Telephone (610) 491-4111

AIR /  
 PRODUCTS 

3 October 1985

Mr. Clayton Richardson  
 Dept. Civil Engineering  
 Marston Hall  
 University of Massachusetts  
 Amherst, MA 01003

Dear Mr. Richardson:

I am pleased to provide you with information regarding the A/O<sup>®</sup> Process for use in your study project at the University of Massachusetts. The two case histories you present provide good examples of the application of the A/O process in retrofitting existing activated sludge plants. As you know, we have had substantial prior experience in this area with very satisfactory results.

From your case history write-up, I have extracted the necessary design information and then prepared a rough cost estimate for conversion to A/O. The results are discussed below.

Palmer Wastewater Treatment Plant

This diffused air type plant will require only minor modifications to the aerator tank. All other process components including the primary and secondary clarifiers and sludge processing equipment is compatible with the A/O process. The influent characteristics to the A/O plant (primary effluent) were assumed as follows:

BOD	180	mg/l
TSS	100	mg/l
P	5	mg/l
FLOW	5.6	MGD

The effluent permit allows discharge of 30/30/11/BOD/TSS/P. Required modifications to the aeration basin will be the addition of a dividing wall to separate the anaerobic zone and the installation of small submerged mixers in the anaerobic zone as depicted in the attached sketch. The cost of conversion is estimated as follows:



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3 October 1985

1. Staging Wall Construction	\$ 20,000
2. Submersible Mixers (8)	\$ 40,000
3. A/O Technology Fee (\$80,000/MGD)	<u>\$448,000</u>
TOTAL COST	\$508,000

In addition, removal of the static mixer/aerators from the anaerobic zone is necessary.

#### North Attleborough

North Attleborough is a surface aeration plant that can be readily converted to the A/O process. The requirement for summer nitrification can be met in the remaining aeration tanks after conversion of one tank to the anaerobic zone. I calculate an F/M of 0.32 day<sup>-1</sup> which meets the requirement for nitrification.

The design influent to the A/O process is estimated to be:

BCO	150
TSS	120
P	10
N	10

The summer effluent requirements are 5/7/1/1//BOD/TSS/P/N.

The mechanical requirements of conversion are the addition of staging walls to the first tank of the aeration basin and installation of mixers in this zone. Because of the tank configuration only four submerged mixers will be required rather than the eight needed for the Palmer Plant.

I estimate the cost of conversion to the A/O process for North Attleborough to be:

1. Construction of Staging Walls	\$ 10,000
2. Submersible Mixers (4)	\$ 20,000
3. A/O Technology Fee (\$80,000/MGD)	<u>\$448,000</u>
TOTAL COST	\$478,000

An additional benefit to be expected from the A/O process would be an improvement in secondary sludge concentration to about 10% solids.

One note of caution for this plant: supplemental alum or ferrous sulfate addition will probably be required due to the unusually high influent phosphorus. A primary effluent phosphorus of 8 mg/l or less is assumed normal while this plant should be around 10 mg/l given the 14 mg/l in the raw influent.

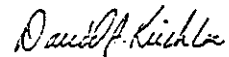
page 3  
3 October 1985

Please consider the proposed tank modifications for the two plants to be tentative; a complete study of structures would be necessary to provide final designs and this is beyond the scope of this project.

I hope this information will be of help in completing your research project. If we can provide any additional information please call.

Best wishes for a successful professional career.

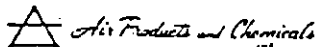
Sincerely yours,



David J. Krichten

DJK/der  
Attachment

FORM NO. 8100

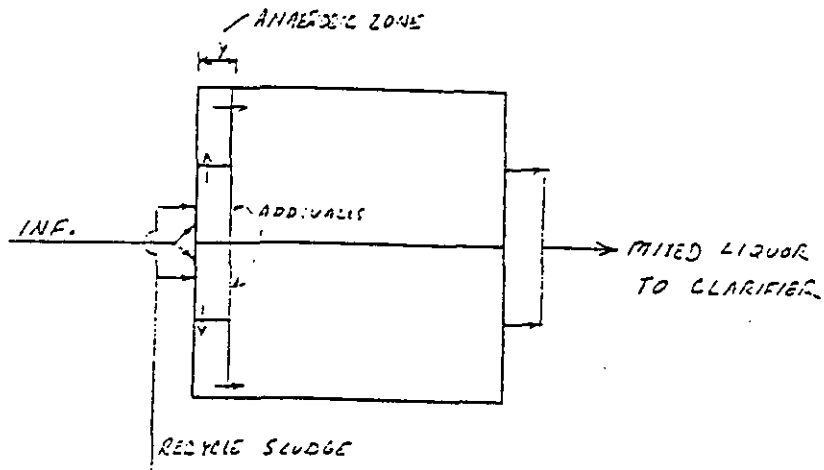


BY DJA DATE 10-2-81

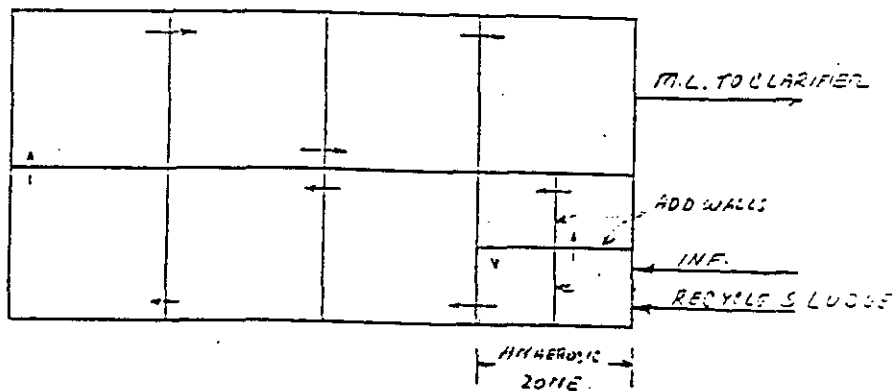
SHEET NO. \_\_\_\_\_ OF \_\_\_\_\_

SUBJECT TANK MODIFICATIONS FOR RETROFIT OF A/D PROCESS

JOB NO. \_\_\_\_\_



PALMEL STP AERATION TANK RETROFIT TO A/D  
(ONE OF TWO TRAINS)



NORTH ATTELBOROUGH STP AERATION TANK RETROFIT

NOTE: ARROWS INDICATE  
DIRECTION OF FLOW

A P P E N D I X 3

CHEMICAL COSTS DETERMINATIONS

## PALMER:

## Phostrip Addition:

$200 \text{ mg/l} \times 8.34 = 1668 \text{ lb, million gallons}$   
 $1668 \text{ lb, mil. gal.} \times .13 \times 3.5 \text{ mgd} = 759 \text{ lb, day}$   
 $759 \text{ lb, day} \times \$6.00, 100 \text{ lbs} = \$45.54, \text{ day}$   
 $45.54 \times 153 \text{ day, yr} = \$7,000., \text{ yr}$

## Conventional Precipitaion:

reduce from 6 mg/l to 1 mg/l = 85% removal  
 Alum 16:1 - 16(6) = 96 mg/l  
 $96 \text{ mg/l} \times 8.34 = 800 \text{ lb, million gallons}$   
 $800 \text{ lb, mil. gal.} \times 3.5 \text{ mgd} = 2800 \text{ lb, day}$   
 $2800, .485 \times \$12.00, 100 \text{ lbs} = \$693, \text{ day}$   
 $693 \times 153 \text{ day, yr} = \$106,000, \text{ yr}$

## NORTH ATTLEBORO:

## Phostrip:

$200 \text{ mg/l} \times 8.34 = 1668 \text{ lb, million gallons}$   
 $1668 \text{ lb, mil. gal.} \times 2.4 \text{ mgd} \times 0.13 = 520 \text{ lb, day}$   
 $520 \text{ lb, day} \times \$6.00, 100 \text{ lbs} = \$31.20, \text{ day}$   
 $122 \text{ day, yr} \times 31.20 = \$3,800., \text{ yr}$

A<sub>2</sub>O:

$3 \text{ mg/l} \times 16 \text{ lb, mg/l} = 48 \text{ mg/l} \times 2.5 \text{ mgd} = 120 \text{ lb, mil gal}$   
 $120 \times 8.34 = 1000 \text{ lb, day}$   
 $1000, .485 \times \$12.00, 100 \text{ lbs} = \$247.00, \text{ day}$   
 $122 \text{ day, yr} \times 247 = \$30,200, \text{ yr}$

## Present Method (Chemical Precipitation):

$75 \text{ gal, day liq. alum} \times 11.2 \text{ lbs, gallon} \times \$12.00, 100 \text{ lbs} =$   
 $\$101.00, \text{ day}$   
 $120 \text{ gallons, day FeSO}_4, 7.48 \text{ gal, ft}^3 \times 65 \text{ lb, ft}^3 =$   
 $1045 \text{ lb, day, .75} = 1393 \text{ lb, day}$   
 $1393 \times \$25.00, 100 \text{ lbs} = \$350.00, \text{ day}$

total cost = (101 + 350) x 122 day, yr = \$55,000, yr