TECHNOLOGY TRANSFER SEMINAR
on
BIOLOGICAL PHOSPHORUS REMOVAL
IN MUNICIPAL
WASTEWATER TREATMENT

Wednesday, 17 April
and
Thursday, 18 April, 1985

DELTA LAKESIDE HOTEL
Penticton, British Columbia

Sponsored by:

Environmental Protection Service,
Environment Canada

in conjunction with
British Columbia Water and Waste Association
1985 "Clear Water" Conference
14 to 16 April 1985

B.C.W.W.A. PENTICTON '85
APRIL 14, 15, 16
EPS TECHNOLOGY TRANSFER
APRIL 17, 18

▌ Environment	Environnement
Canada	Canada
Environmental	Service de la
Protection	protection de
Service	l'environnement

PROCEEDINGS OF THE SEMINAR

ON

BIOLOGICAL PHOSPHORUS REMOVAL

IN

MUNICIPAL WASTEWATER TREATMENT

PENTICTON, BRITISH COLUMBIA

APRIL 17 AND 18, 1985

PROGRAM

Wednesday, 17 April

8:00 a.m. -   **Registration** (Hotel lobby)

9:00 a.m. -   **Background to Biological Phosphorus Removal**
H. Melcer, Ph.D., P.Eng.
Environment Canada, Wastewater
Technology Centre,
Burlington, Ontario.

9:30 a.m. -   **Introduction to Biological Phosphorus Removal**
B.H. Paepcke, M.A.Sc.
Environment Canada, Wastewater
Technology Centre,
Burlington, Ontario.

10:15 a.m. -   **Coffee Break**

10:45 a.m. -   **Introduction to Biological Phosphorus Removal**
(continued)

12:15 p.m. -   **Lunch Break**

1:30 p.m. -   **Introduction to Biological Phosphorus Removal**
(continued)

3:00 p.m. -   **Coffee Break**

3:30 p.m. -   **Technical and Economic Feasibility of Retrofitting Existing Municipal Wastewater Treatment Plants in Canada for Biological Phosphorus Removal**
S.G. Nutt, M.Eng., P.Eng.
Canviro Consultants Ltd.,
Kitchener, Ontario

4:15 p.m. -   **Retrofit of a Full Municipal Treatment Plant at Pontiac, Michigan for Biological Phosphorus Removal**
S.J. Kang, Ph.D., P.E.
McNamee Porter & Seeley, Ann Arbor, Michigan

PROGRAM

Thursday, 18 April

8:30 a.m. - **Background to the Selection of the Modified Bardenpho Process for Kelowna**
City of Kelowna

8:50 a.m. - **Design of the Kelowna Pollution Control Centre**
P.J. Leslie, M.Eng., P.Eng.
Knight & Piésold Ltd., Vancouver

9:30 a.m. - **Operating Experiences with the Kelowna Pollution Control Centre**
W.K. Oldham, Ph.D., P.Eng.
University of British Columbia
G.M. Stevens, C.E.T.
City of Kelowna

10:15 a.m. - **Coffee Break**

10:30 a.m. - **Operating Experiences with the Kelowna Pollution Control Centre** (continued)

11:30 a.m. - **Lunch Break**

12:30 p.m. - **Depart by bus and private vehicles to Kelowna Pollution Control Centre**

1:30 p.m. - **Tour of Kelowna Pollution Control Centre**

4:30 p.m. - **Bus departs for Kelowna Airport and return to Penticton**

For further information contact:

Tom Tevendale, EPS,
West Vancouver, B.C.   (604)666-6711
Bruce Paepcke, EPS,
Burlington, Ontario   (416)637-4741
Venue West Management Services,
Vancouver B.C.   (604)681-5226

BACKGROUND TO BIOLOGICAL PHOSPHORUS REMOVAL

Henryk Melcer

Environment Canada

Wastewater Technology Center

Burlington, Ontario

BACKGROUND TO BIOLOGICAL PHOSPHORUS REMOVAL

Henryk Melcer
Environment Canada
Wastewater Technology Centre
Burlington, Ontario L7R 4A6

ABSTRACT

The need for nutrient control in wastewater treatment plants, specifically phosphorus removal is discussed. Phosphorus removal is achieved in Canada primarily by chemical precipitation. Specific examples are presented. Biological phosphorus removal in Canada is reviewed and the future of biological phosphorus removal technology is assessed.

INTRODUCTION

Intensive agricultural practice and large-scale urbanization has, in some areas, led to a rapid increase in nutrient loading to receiving water courses and an increase in their eutrophic state. Eutrophication can cause increases in algal and weed populations which create algal scums, odours and a reduction in the clarity and dissolved oxygen concentration of stream and lake waters. These effects can interfere with using these waters for drinking and recreational purposes. Effluents from wastewater treatment plants have been identified as major contributors to eutrophication. Nutrient removal from treatment plant effluents would assist in alleviating this problem. In natural waters, algae require a nitrogen to phosphorus ratio of 15:1 but since this ratio is approximately 4:1 in treatment plant effluents, it would appear that nitrogen should be the limiting nutrient. However, due to the nitrogen-fixing capability of various algal species, that is, their ability to take up atmospheric nitrogen, phosphorus has been shown to be the limiting nutrient in most cases.

Other sources of phosphorus that contribute to nutrient loads are diffuse in nature, for example, surface run off, fertilizer applications and, to a lesser extent, wild and domestic animal wastes and atmospheric precipitation. Limiting phosphorus emissions via diffuse source control or the control of point sources such as treatment plant discharges has been debated extensively in the literature in view of the considerable cost of either option.

In the early 1970's, bans on phosphates in detergents were considered to be a relatively quick and easy way to reduce one of the major sources of phosphorus in municipal wastewater (Flynn, 1982). Canada enacted nationwide controls on the use of phosphorus in detergents, reducing levels to 2.2% by weight by 1972. Five of the U.S. Great Lakes States (Indiana, New York, Michigan, Minnesota, & Wisconsin) followed suit. However, the success of this campaign has been contested by the detergent industry who have published data which claim that the elimination of detergent phosphorus has not measurably improved receiving water quality (Wendt, 1982; Maki et al., 1984).

It is evident from detailed cost-effectiveness studies (Chapra et al., 1983) that the least-cost strategy for phosphorus management relies on a combination of both point and diffuse source controls and zoned, rather than uniform, treatment. Also, the rate of treatment costs escalates rapidly as water quality objectives become more stringent. Analysis of phosphorus management strategies are, therefore, site specific.

The present paper addresses phosphorus control in point source discharges, specifically those from municipal wastewater treatment plants. In a typical activated sludge plant, normal microbial growth will account for a 10 to 30% reduction in influent phosphorus depending upon the value of solids retention time used and 1 mg/L phosphorus is removed for every 100 mg/L BOD_5 removed. Under some operating conditions it has been observed that activated sludge can remove more phosphorus than is required for normal growth. The terms "luxury uptake" and "enhanced phosphorus removal" were introduced to describe this phenomenon which became the basis of the three major commercial variations of biological phosphorus removal technology that exist at this time.

PHOSPHORUS REMOVAL IN CANADIAN WASTEWATER TREATMENT PLANTS

The major areas in Canada where phosphorus control is practised are located in the Great Lakes Basin and in Western Canada, that is, in the Okanagan, Thompson and Fraser River Valleys of British Columbia, and, to a lesser extent, in the rain shadow of the Rockies in Alberta. Nutrient control has been stimulated by the need to correct the eutrophied state of some receiving lake waters and by communities in which the capacity of receiving waters to accept treated sewage effluents is often reduced by inadequate dilution during seasonal low flow periods.

There are a number of alternatives available for the control of phosphorus in wastewaters, namely, chemical precipitation, biological-chemical techniques and purely biological methods. Chemical precipitation is normally used in existing plants by chemical addition to the plant effluent. In biological-chemical systems, the ability to remove phosphorus biologically is complemented by the addition of chemicals. Biological processes achieve phosphorus removal without the addition of chemicals.

Chemical Addition

In 1969, the International Joint Commission (IJC) recommended that all treatment plant discharges to the Lower Great Lakes Basin (Lakes Erie and Ontario) meet an effluent standard of 1.0 mg/L total phosphorus. This recommendation was formalized in the Canada-U.S. Great Lakes Water Quality Agreement of 1972. The Canada-Ontario Agreement of 1971 accelerated pollution control programs in the Lower Great Lakes Basin to meet the IJC recommendations. A $250 million capital works program upgraded sewage collection systems and treatment works. Of this, approximately $40 million funded the installation of phosphorus removal equipment. An additional $6 million were devoted to research in phosphorus removal technology and treatability studies which led to the integration of chemical precipitation processes at existing treatment plants. Investigations by both the Ontario Ministry of Environment and Environment Canada's Wastewater Technology Centre ensured the successful implementation of this phosphorus removal technology at most Southern Ontario plants by the mid-1970s. A two-stage program was adopted; initial jar testing studies were followed by long term

full scale studies. This yielded information for chemical and dosage selection as well as the optimum point of chemical application and compatibility with the existing waste treatment process type and operation. The approach most frequently used is the addition of iron or aluminum salts to the wastewater at some point in the conventional secondary treatment system. The number of delinquent plants has been progressively reduced, as reported for 1982 by the Great Lakes Water Quality Board (1983). Of the larger Canadian treatment plants in the Lower Great Lakes Basin (45 400 m^3/d (10 x 10^6 IGPD)), only Hamilton and Toronto Humber (both at 1.2 mg/L total phosphorus) exceeded the effluent phosphorus standards.

Very low effluent concentrations may be achieved by dual point alum addition. For example, in the Stratford-Avon River Environmental Management Project in Ontario, an average effluent total phosphorus of 0.1 mg/L was consistently achieved by adding 68 mg/L of alum to the aeration basin effluent and a secondary dose of 15 mg/L in advance of the effluent granular media filter (Zukovs and Engler, 1982) However, filter backwash requirements nearly doubled, resulting in a 2.3 fold increase in solids load returned to the primary clarifiers.

In some Ontario sewage treatment plants, effluent is stored in lagoons when the capacity of receiving waters is limited by inadequate dilution during low flow periods in the summer. On the other hand, the kinetics of a river system may dictate a different priority. For example, in Kamloops, British Columbia, the Thompson River is at its lowest flow in winter and therefore a minimal phosphorus discharge is required during the winter months. Treatment consists of a series of four lagoons, the last of which is drawn down in the fall and used as a reservoir in the winter for the effluent from the previous three lagoons. During the summer, alum addition lowers the effluent total phosphorus levels to 0.8 mg/L (Shepherd and Vachon, 1981).

Biological Phosphorus Removal

At the present time, there is only one full-scale Canadian biological phosphorus removal facility in Kelowna, B.C. It was built in 1981 in response to a need to remove both nitrogen and phosphorus. To meet these requirements a modified Bardenpho process was installed. Environment Canada sponsored the University of British Columbia to monitor the performance of the plant, to determine the viability of using oxidation-reduction potential (ORP) probes as a monitoring tool and to evaluate a number of control strategies to maximize phosphorus removal. Performance data since startup have shown that high quality effluent can be produced (Oldham, 1984).

RESEARCH

Much of the early research and development of biological phosphorus removal and the subsequent full-scale application of the technology were undertaken in South Africa in the 1970's. Subsequently, the technology has been applied, and further refined, in the U.S., Denmark, France, the Netherlands, West Germany and Japan. In Canada, limited research work is in progress on the fundamentals of biological phosphorus removal. The two major centres of research are at the Universities of British Columbia and

Toronto. Specifically, work at the University of British Columbia is focussed on:

- biochemistry of polyhydroxybutyrate storage during biological phosphorus removal
- effect of different sewage characteristics and specific substrates on phosphorus release and uptake
- effect of pH, temperature and solids retention time on VFA production in primary sludge
- viability of oxidation-reduction potential (ORP) monitoring of biological phosphorus removal

At the University of Toronto, work is directed at:

- specific substrate induction of phosphorus release
- optimization of acetate production in sludge digestion
- electrical stimulation of phosphorus release

The Ontario Ministry of Environment is investigating the degree of phosphorus removal in artificial marsh treatment systems (Reed, et al., 1984; Black, 1984).

BIOLOGICAL PHOSPHORUS REMOVAL IN THE FUTURE

In a further effort to reduce phosphorus inputs to the Great Lakes Basin, the 1978 Canada-U.S. Great Lakes Water Quality Agreement recommended a proposed standard that called for all plants discharging more than 4540 m^3/d (1.0 x 10^6 IGPD) in the Upper Great Lakes Basin (Lakes Superior, Michigan and Huron) to achieve a maximum effluent total phosphorus concentration of 1.0 mg/L. For plants in the Lower Great Lakes Basin (Lake Ontario and Erie), an effluent total phosphorus concentration of 0.5 mg/L was considered.

Using chemical precipitation technology, low effluent total phosphorus levels can generally only be achieved with high chemical dosage rates, resulting in increased sludge production, and by providing additional clarifier surface area and/or effluent filtration. The capital and operating costs of these additional facilities are significant as are the increased sludge handling costs. The potential for such large expenditures has stimulated interest in alternative technologies such as biological phosphorus removal. Also, these high incremental costs led authorities to query the proportion of the emitted phosphorus which is bioavailable. Accordingly, a program was carried out at the Wastewater Technology Centre to provide additional technical information on the bioavailability of phosphorus in municipal wastewater and the effect of wastewater treatment on bioavailable phosphorus (BAP). Table 1 summarizes the effluent phosphorus characteristics for the nine plants that were evaluated in Southern Ontario (Canviro, 1983).

The lowest effluent total phosphorus (0.1 mg/L) and BAP concentrations were produced by the dual-stage chemical addition system tested at the Stratford treatment plant. For all the treatment processes monitored, with the exception of the Listowel artificial marsh system which exhibited significantly lower phosphorus removal during winter, there were no measured differences between phosphorus and BAP removal characteristics under winter or summer conditions.

TABLE 1. EFFLUENT CHARACTERISTICS - BAP STUDY

Process	Effluent Total P (mg/L)	Average Per Cent of Total P that is Bioavailable
Continuous Discharge Lagoon without P-Removal	1.9	90
Secondary Plant with P-Removal	3.1	91
Primary Stage of Conventional Activated Sludge Plant	2.0 to 4.0	64
Secondary Plant with P-Removal	0.3 to 1.3	52
Secondary Plant with P-Removal and Filtration	0.1 to 0.7	49
Seasonal Discharge Lagoon with Batch Alum Treatment	0.8	19
Artificial Marsh Treating Lagoon Effluent	1.3	54

The bioavailability of the soluble fraction tended to dominate the total BAP of secondary and filtered effluents, usually representing from 65 to more than 85 per cent of the total BAP. The proportion of total phosphorus that was bioavailable remained at a relatively high level even after tertiary filtration. This was for effluents with average suspended solids concentration of 15 mg/L or less. The study concluded that additional phosphorus control should be focussed on the control of the soluble (mainly ortho-phosphate) fraction of effluents.

It is evident that the established phosphorus removal technology, chemical precipitation, can achieve high levels of BAP removal. Consequently, biological phosphorus removal technology faces a number of challenges, that is, to produce, reliably, at low water temperatures, an effluent containing a minimum total phosphorus concentration of 0.5 to 1.0 mg/L that has minimum biological availability. In view of the number of incumbent chemical precipitation schemes, retrofitting existing facilities with biological phosphorus removal technology has to be economically attractive to potential users. Biological phosphorus removal has to be demonstrated to be at least as cost effective as chemical precipitation.

REFERENCES

Black, S.A. (1984). The Use of Marshlands in Wastewater Treatment - A Novel Concept. Ontario Ministry of Environment.

Canviro Consultants Ltd. (1983). Investigation of Bioavailable Phosphorus (BAP) in Municipal Wastewaters and the Effect of Wastewater Treatment on BAP. Wastewater Technology Centre, Environment Canada.

Chapra, S.C., et al. (1983). Effectiveness of Treatment to Meet Phosphorus Objectives in the Great Lakes. J. Water Pollut. Control Fed., 55(1), 81-91.

Flynn, K.C. (1982). New Challenges in the Great Lakes States to Banning Phosphorus in Detergents. J. Water Pollut. Control Fed., 54(10), 1342-5.

Great Lakes Water Quality Board (1983). 1983 Report on Great Lakes Water Quality. International Joint Commission, Windsor, Ontario.

Maki, A.E., et al. (1984). The Impact of Detergent Phosphorus Bans on Receiving Water Quality. Water Res., 18(7), 893-903.

Oldham, W.K. (1984). Full-scale Optimization of Biological Phosphorus Removal at Kelowna, Canada. Presented at IAWPRC Post-conference Seminar on Enhanced Biological Phosphorus Removal from Wastewater, Paris, France.

Reed, S. et al. (1984). Wetlands for Wastewater Treatment in Cold Climates. Presented at Water Reuse Symposium III, San Diego, CA.

Shepherd, R.B. and D.T. Vachon (1981). An Investigation of the Kamloops Sewage Lagoons. EPS Pacific Regional Report 82-101, Environment Canada.

Wendt, R.H. Discussion of: A Preliminary Assessment of Michigan's Phosphorus Detergent Ban. J. Water Pollut. Control Fed., 54(10), 1425-7.

Zukovs, G. and F. Engler (1982). Urban Phosphorus Management - the Stratford Experience. Presented at the PCAO Conference, Toronto, Ontario.

INTRODUCTION TO BIOLOGICAL PHOSPHORUS REMOVAL

Bruce H. Paepcke

Environment Canada

Wastewater Technology Center

Burlington, Ontario

INTRODUCTION TO BIOLOGICAL PHOSPHORUS REMOVAL

Bruce H. Paepcke
Wastewater Technology Centre
Environmental Protection Service
Environment Canada
Burlington, Ontario L7R 4A6

ABSTRACT

After a brief review of the early history of biological phosphorus removal, the development of the Modified Bardenpho process in South Africa is described. The principles of nitrification and denitrification are discussed along with the general biochemistry of biological phosphorus removal and the variables which have an impact on biological phosphorus removal. The application of the Modified Bardenpho process at full-scale in South Africa and the United States is described. The main characteristics of the UCT, Pinetown, A/O, PhoStrip and Biodenipho processes are given along with the full-scale operating results for these processes.

EARLY HISTORY OF BIOLOGICAL PHOSPHORUS REMOVAL

Levin and Shapiro (1965) were among the first researchers to study the phenomenon of enhanced biological phosphorus removal in the activated sludge process, i.e. phosphorus removal in amounts greater than the normal metabolic requirements of the activated sludge micro-organisms. In laboratory and full-scale experiments at the District of Columbia Sewage Treatment Plant, they found that phosphorus removal as high as 80 percent could be achieved without the addition of chemicals by using aeration rates three to four times greater than those normally used. The economics of doing this on a long term basis were not considered. They described the uptake and storage of phosphorus by the sludge micro-organisms as "luxury uptake", and hypothesized that the capacity of the micro-organisms to take up phosphorus could be increased if the phosphorus content of the micro-organisms was regularly depleted. They had observed in laboratory experiments that phosphorus was secreted from the micro-organisms under acidic conditions, and proposed that the phosphorus could be "stripped" from the micro-organisms by exposing the return sludge to acidic conditions created by means of pH adjustment.

Several years later, Shapiro et al. (1967) observed at the Baltimore Sewage Treatment Plant that phosphorus which was taken up in excess quantities by the sludge biomass in the aeration basins was later released into solution in the bottom of the secondary clarifiers under conditions of low or zero dissolved oxygen and low redox potential. They proposed that the phosphorus content of the sludge could be deliberately depleted by exposure to this type of condition prior to return to the aeration basins rather than by exposure to an acidic environment. This work led to the development of the PhoStrip process (Levin et al., 1972; Levin et al., 1975) which will be discussed in more detail later in this paper.

Phosphorus removal at levels much greater than that normally experienced was observed at a number of other activated sludge plants in the U.S. in the late 1960's. At the Rilling Plant in San Antonio, Texas (Vacker et al., 1967), phosphorus removal as high as 96 percent without the addition of chemicals was observed. Two other plants of similar design treated the same wastewater at the same site, but such an abnormally high level of phosphorus removal was not observed in these. The authors considered that the differences in phosphorus removal between the plants were due to differences in the operation of the plants, and recommended that for maximum phosphorus removal the F/M ratio be 0.5 kg BOD/kg MLSS·d, the SVI be maintained at 150 mL/g, and the DO concentration be maintained at 2.0 mg/L at the midpoint of the aeration basin and 5.0 mg/L at the discharge end of the aeration basin. They noted that the front half of the aeration tanks in the Rilling Plant was depleted of oxygen due to clogged diffuser tubes, and measured soluble orthophosphate concentrations in the front half of the aeration tanks much greater than that in the influent wastewater, but did not consider these observations to be related to the high level of phosphorus removal that was achieved.

Phosphorus removal greater than 95 percent was observed at the Hyperion Treatment Plant in Los Angeles during a six month period in 1969 (Bargman et al., 1971). The investigators considered that in order to achieve such high removal efficiencies in an activated sludge plant, it was necessary to provide much higher aeration rates than normal, design for nitrification in order to ensure that carbonaceous oxygen demand was completed, and make certain that phosphorus removed from the process in the waste activated sludge not be returned to the process in supernatants or any other way.

Milbury et al. (1971) carried out a study at the Back River Wastewater Treatment Plant in Baltimore, Maryland after having observed phosphorus removal of 82 percent without the addition of chemicals. They concluded that the following operating conditions appeared to be necessary for enhanced phosphorus removal:

1. Prevention of recycle of phosphorus in the waste activated sludge back to the treatment process.
2. Operation of aeration basin in plug flow configuration.
3. Maintaining aeration so that the D.O. concentration was not limiting.
4. Wasting of activated sludge on a continuous rather than intermittent basis.
5. Maintaining MLSS concentration at 1200 or 1300 mg/L or greater.
6. Maintaining aerobic conditions in the secondary clarifier in order to prevent the release of phosphorus into the effluent.

The authors noted that in this plant and all the other full-scale activated sludge plants in the U.S. where unusually high levels of phosphorus removal had been recorded, the hydraulic configuration of the aeration basin was plug flow and there was a release of phosphorus from the sludge micro-organisms near the inlet end of the aeration basin followed by an uptake of the phosphorus along the length of the aeration basin. This observation led to the above conclusion that plug flow was important, but the common release of phosphorus at the inlet end of the aeration basin was not discussed further.

DEVELOPMENT OF THE MODIFIED BARDENPHO PROCESS

Apart from the development of the PhoStrip process in the U.S. previously referred to, most of the early research and development work on biological phosphorus removal took place in South Africa in the early 1970's. Eutrophication, i.e. algal blooms due to excessive nutrient inputs, began to be noticeable in some South African lakes and impoundments in the late 1960's and early 1970's. This stimulated some interest in treatment processes for the removal of nitrogen and phosphorus from wastewater discharges. In 1971, Dr. James Barnard returned to South Africa after having spent some years studying at U.S. universities, and rejoined the National Institute for Water Research (NIWR) of the Council for Scientific and Industrial Research (CSIR). One of his initial tasks there was to conduct an economic evaluation of the various nutrient removal processes available at the time. In doing this he first evaluated the processes from a technical point of view, and the results of this led him to question the feasibility of applying some of these processes on a large scale in South Africa.

One of the first processes that Barnard looked at was the three stage separate sludge system for carbon and nitrogen removal tested at pilot-scale by Barth et al. (1968). As shown in Fig. 1, three completely separate reactor/clarifier units were used in series for carbon removal, nitrification and denitrification respectively. In this system, carbon removal and nitrification took place in separate reactors so that they could be individually optimized, but they of course can occur together in the same tank as in a nitrifying extended aeration plant.
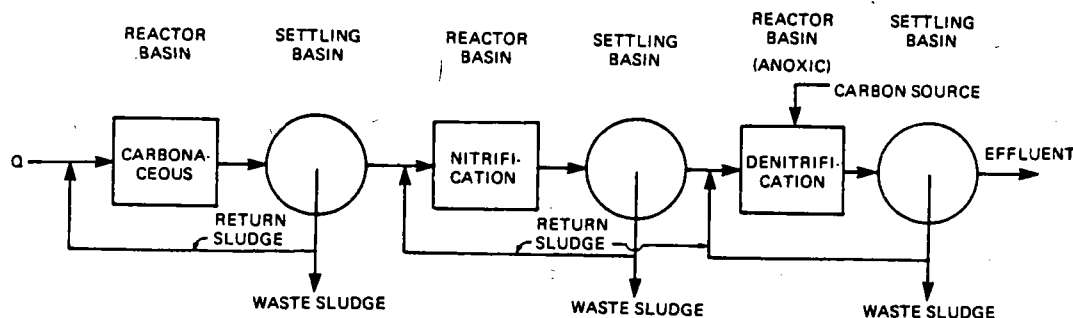

FIG. 1.   THREE SLUDGE SYSTEM FOR CARBON AND NITROGEN REMOVAL
(From Water Pollution Control Federation, 1977)

A review of the principles of nitrification and denitrification will be useful at this point as most of the processes for biological phosphorus removal incorporate biological nitrogen removal, and nitrogen removal in these processes can have an impact on phosphorus removal.

Nitrification

Nitrification is the conversion of ammonia nitrogen (NH_4^+) to nitrate nitrogen (NO_3^-). In this reaction, nitrogen is not removed from the waste stream but simply converted from one form to another. It is a two step

reaction. In the first step, an autotrophic bacteria known as <u>Nitrosomonas</u> obtains its energy for growth by oxidizing ammonia nitrogen to nitrite nitrogen (NO_2^-) (Grady and Lim, 1980). This reaction is complex and not yet completely understood, but can be represented by the following simplified equation (Ekama and Marais, 1984a).

$$NH_4^+ + 1.5\ O_2 \xrightarrow{\text{Nitrosomonas}} NO_2^- + H_2O + 2H^+ \qquad (Eq.\ 1)$$
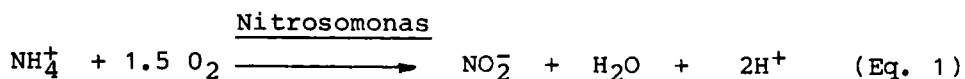
In the second step, another autotrophic bacteria known as <u>Nitrobacter</u> obtains its energy for growth by oxidizing nitrite nitrogen to nitrate nitrogen as shown in the following equation.

$$NO_2^- + 0.5\ O_2 \xrightarrow{\text{Nitrobacter}} NO_3^- \qquad (Eq.\ 2)$$
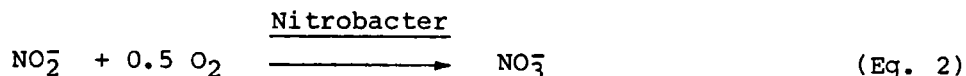
The overall reaction can be represented by the following simplified equation.

$$NH_4^+ + 2O_2 \xrightarrow{\substack{\text{Nitrifying}\\\text{bacteria}}} NO_3^- + H_2O + 2H^+ \qquad (Eq.\ 3)$$
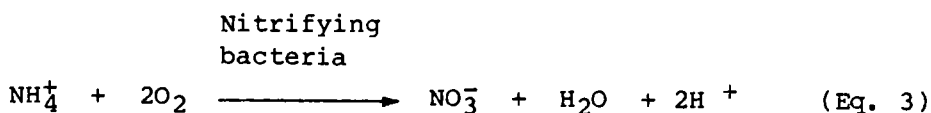
These nitrifying bacteria have relatively low growth rates even at warm temperatures, and as a result, it is necessary to use a relatively long solids retention time (SRT) in an activated sludge process in order to allow time for a buildup in their population to occur and be maintained. In the temperate parts of Canada, the SRT required to maintain full nitrification during the winter is 15 to 30 d depending on local conditions. On the other hand, in the summer months some nitrification will occur at very low SRT's of 2 or 3 d, although to ensure full nitrification a longer SRT of at least 10 d should be used.

Equation 3 indicates that two moles of O_2 are required for conversion of one mole of NH_4^+ to NO_3^-. This amount of oxygen is in addition to that required for oxidation of carbonaceous matter in the activated sludge process, and must be made available for nitrification to occur. On a mass basis, the oxygen requirement for nitrification is 4.57 mg O_2 per mg NH_4^+ as N converted (Ekama and Marais, 1984a).
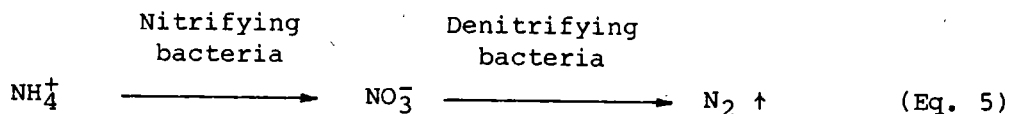
Finally, nitrification can be inhibited by a variety of metals and organic compounds, and these must be absent for nitrification to occur (Grady and Lim, 1980).

Denitrification

As mentioned earlier, during nitrification nitrogen is not removed from the waste stream but simply converted from the ammonia form to the nitrate form. Nitrate, however, can be converted to nitrogen gas (N_2) by many different types of bacteria which, in the absence of free oxygen, use nitrate in place of oxygen as a final hydrogen acceptor in microbial respiration (Grady and Lim, 1980). This reaction can be represented by the following equation.

$$\text{NO}_3^- \xrightarrow{\substack{\text{Denitrifying} \\ \text{bacteria}}} \text{N}_2\uparrow \qquad\qquad \text{(Eq. 4)}$$

The nitrogen gas produced in this reaction is released to the atmosphere from the treatment process, and this is the actual means of nitrogen removal from the system. Thus the overall process for nitrogen removal in an activated sludge process can be represented by the following simplified equation.

$$\text{NH}_4^+ \xrightarrow{\substack{\text{Nitrifying} \\ \text{bacteria}}} \text{NO}_3^- \xrightarrow{\substack{\text{Denitrifying} \\ \text{bacteria}}} \text{N}_2\uparrow \qquad\qquad \text{(Eq. 5)}$$

In order for denitrification to occur in the third reactor shown in Fig. 1 or any other denitrification reactor, the dissolved oxygen concentration in the reactor must be near zero. Thus air is not added to the reactor and the mixed liquor is kept in suspension by a simple mechanical mixing system, such as a stirrer. This condition where dissolved oxygen is absent but nitrate is present is known as anoxic (Ekama et al., 1984).

Carbon substrate is required by the denitrifying bacteria as an energy source. Stoichiometrically, 8.6 mg of COD are required for denitrification of 1.0 mg of NO_3^- as N (Ekama and Marais, 1984b). In the system shown in Fig. 1, almost all of the available carbon was removed in the first stage, and as a result, methanol was added to the denitrification reactor as an external carbon source. However, methanol can also be used as a fuel, and the cost of adding it to a full-scale plant as a carbon source can be considerable. This caused Barnard to consider alternate systems for nitrogen removal.

Ludzack-Ettinger Process

One of the other nitrogen removal processes that Barnard looked at was the "semi-aerobic" activated sludge process proposed by Ludzack and Ettinger (1962). They performed laboratory studies using an aeration basin with a small unaerated anoxic zone at the inlet end as shown in Fig. 2. Nitrification occurred in the aerated zone and nitrified mixed liquor was recycled from there back to the anoxic zone by the mixing action provided by the aeration system. A significant amount of nitrogen removal in the anoxic zone by means of denitrification was observed under some conditions. Methanol was not added as a carbon source for denitrification, but rather the carbon in the influent wastewater served as the carbon source for denitrification in the anoxic zone. This was seen as a great advantage. Not only was an external carbon source not necessary, but the aeration requirement was reduced because some of the carbon in the influent wastewater was removed during denitrification in the anoxic zone.

ANOXIC   AEROBIC
REACTOR REACTOR

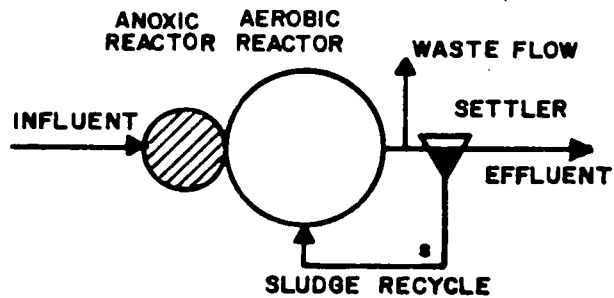WASTE FLOW

SETTLER

INFLUENT

EFFLUENT

SLUDGE RECYCLE

FIG. 2.   LUDZACK-ETTINGER PROCESS
(From Ekama and Marais, 1984b)

Barnard proposed improving the Ludzack-Ettinger process by completely
separating the anoxic and aeration zones into individual basins and
providing an external mixed liquor recycle line from the aeration basin
back to the anoxic basin as shown in Fig. 3.   This modification, which is
sometimes called the Modified Ludzack-Ettinger process, allowed the amount
of nitrate recycled back to the anoxic basin to be directly controlled
rather than relying on recycle by hydraulic backmixing as in the original
configuration.

ANOXIC   AEROBIC
REACTOR   REACTOR

MIXED LIQUOR
RECYCLE a

WASTE FLOW

SETTLER

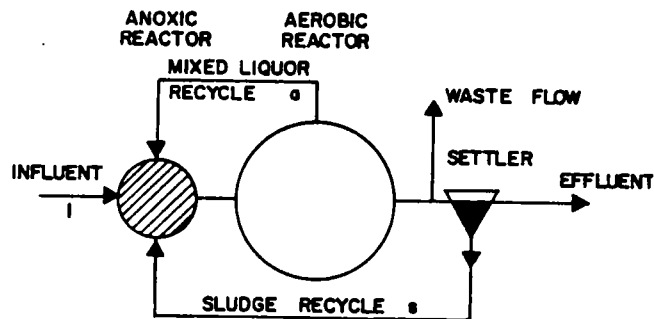INFLUENT

EFFLUENT

SLUDGE RECYCLE s

FIG. 3.   MODIFIED LUDZACK-ETTINGER PROCESS
(From Ekama and Marais, 1984b)

A significant disadvantage of this configuration, however, was that
nitrate produced in the aeration basin that was not recycled back to the
anoxic zone left the system in the plant effluent having never had a chance
to be converted to nitrogen gas.

Wuhrmann Process

Barnard also studied the single sludge system for nitrogen removal
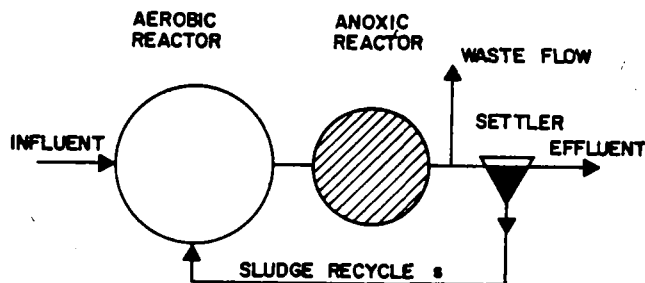proposed by Wuhrmann (1964) shown in Fig. 4.

FIG. 4.   WUHRMANN PROCESS
(From Ekama and Marais, 1984b)

In this configuration, nitrification and carbon removal took place under aerobic conditions in the first basin, and denitrification of the nitrate produced in the first basin occurred under anoxic conditions in the second basin. Due to the fact that almost all of the carbonaceous substrate in the influent wastewater was consumed in the first basin, the micro-organisms in the second basin underwent endogenous respiration, i.e. were in the death phase as opposed to being in a phase of rapid growth. Carbon made available through death and cell lysis under endogenous respiration, however, formed the carbon source for denitrification in the second basin. The major disadvantage of this configuration was that the rate of denitrification was low because the rate of release of carbon under endogenous respiration is quite low, and as a result the second basin had to be very large if significant nitrogen removal was to be achieved.

Bardenpho Process

Barnard proposed compensating for the disadvantages of the Modified Ludzack-Ettinger and Wuhrmann processes by combining the two in the four stage process shown in Fig. 5. He called this configuration the Bardenpho process (Barnard, 1974).
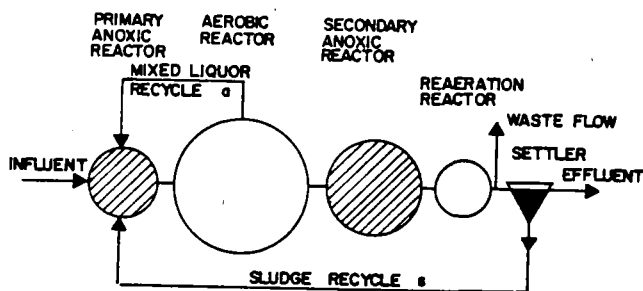

FIG. 5.   BARDENPHO PROCESS
(From Ekama and Marais, 1984b)

Carbon removal and nitrification took place in the main aeration basin. Nitrified mixed liquor was recycled by an external recycle line from this basin back to the primary anoxic basin where, in the absence of free dissolved oxygen, denitrification occurred using the carbon in the influent wastewater as the carbon source. That mixed liquor which was not

recycled back to the primary anoxic basin passed on to the secondary anoxic basin where additional denitrification took place at a slow rate under endogenous respiration. Prior to entering the secondary clarifier, the mixed liquor from the secondary anoxic basin passed through a small reaeration basin for the following reasons:

1. Nitrification of any ammonia released during endogenous respiration in the secondary anoxic basin.

2. To ensure that aerobic conditions existed in the secondary clarifier as any denitrification that occurred there under anoxic conditions would produce nitrogen gas that could cause rising sludge.

3. To ensure that aerobic conditions existed in the secondary clarifier so that any phosphorus that may have been in the sludge was not released into the effluent.

Barnard operated a 100 m^3/d pilot plant using this process to treat settled domestic wastewater starting in October 1972 (Barnard, 1974). Nitrogen removal up to 95 percent was obtained. In addition, phosphorus removal up to 97 percent without the addition of chemicals was observed, particularly when denitrification was complete. During periods when high phosphorus removal was achieved, the soluble orthophosphate concentration in the liquid phase of the mixed liquor varied significantly between the four basins. In particular, the soluble orthophosphate concentration was approximately 30 mg/L in the secondary anoxic basin, which was about three times that in the influent wastewater, while it was about 0.3 mg/L in the reaeration basin. Nitrates were always absent from the secondary anoxic basin. Barnard recognized that phosphorus release was occurring in the secondary anoxic basin of this pilot plant in the absence of free dissolved oxygen and bound oxygen in the form of nitrates, and that the released phosphorus was then removed from solution under aeration in the last basin. He also noted that this was consistent with the observation by Milbury et al. (1971) that in their plant and all the other full-scale activated sludge plants in the U.S. where unusually high levels of phosphorus removal occurred, there was a release of phosphorus from the sludge micro-organisms near the inlet end of the aeration basin followed by an uptake of phosphorus along the length of the aeration basin.

The Phoredox Concept

Based on the above observations, Barnard postulated (Barnard, 1974) that the key requirements for enhanced biological phosphorus removal were that the sludge micro-organisms must first be exposed to an environment free of dissolved oxygen and nitrate under which phosphorus release could occur, and then at a later stage be exposed to an aerated or aerobic environment under which the phosphorus that was previously released into solution and the phosphorus entering in the influent wastewater could either be taken up by the micro-organisms in quantities greater than their normal metabolic requirements, or precipitated from solution as a result of the change in redox potential caused by exposure to the different environments. The condition where free dissolved oxygen and nitrate were absent was referred to as anaerobic. In a later paper (Barnard, 1976), he

proposed the use of a separate basin in which to create the anaerobic conditions necessary for the release of phosphorus from the sludge micro-organisms, and termed this the Phoredox process. This name was derived from "phosphorus" and "redox potential", while the earlier name for his four stage process originally intended for nitrogen removal was based on "Barnard", "denitrification" and "phosphorus removal".

Barnard (1976) proposed applying the Phoredox process to the Bardenpho process, i.e. adding an anaerobic basin in front of the existing four stages as shown in Fig. 6.
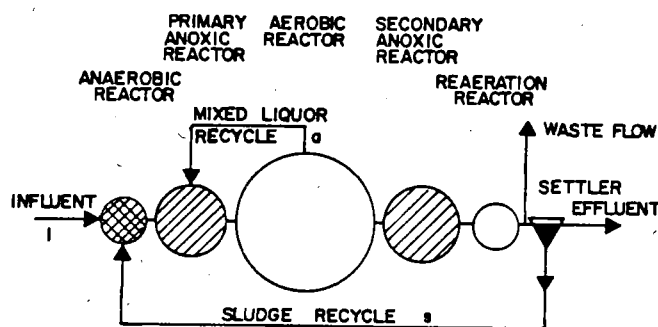

FIG. 6. MODIFIED BARDENPHO PROCESS
(From Ekama et al., 1984)

He also postulated that enhanced biological phosphorus removal could be achieved in a simple activated sludge plant not designed for nitrogen removal by adding an anaerobic basin to the head end of the plant as shown in Fig. 7. In this configuration, nitrification in the aeration basin would have to be minimized so that nitrate was not discharged into the anaerobic basin via the return sludge.
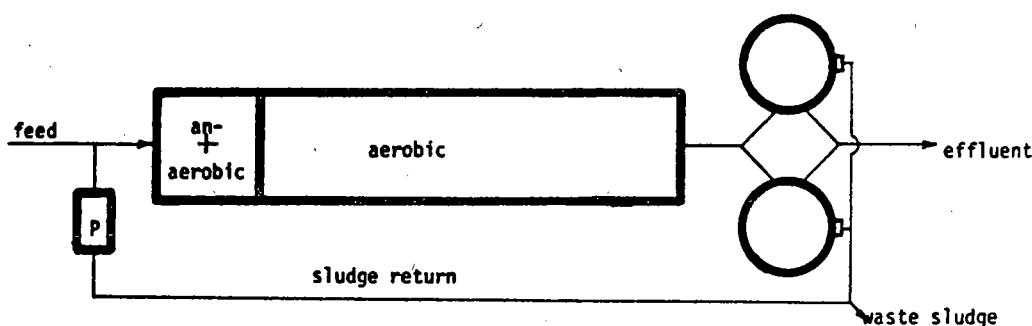

FIG. 7. PHOREDOX PROCESS WITHOUT NITROGEN REMOVAL
(From Barnard, 1976)

Modified Bardenpho Process

The five stage process for biological nitrogen and phosphorus removal shown in Fig. 6 and in another representation in Fig. 8 below is the basic process used at Kelowna and the many full-scale nutrient removal plants in South Africa. It is widely known as the Modified Bardenpho process. To

the best knowledge of the author, the only full-scale plant in the world ever built utilizing the original four stage process shown in Fig. 5 was at a timber mill in South Africa. Often though, one hears of the Bardenpho process, but this usually refers to the five stage process known more formally as the Modified Bardenpho process.
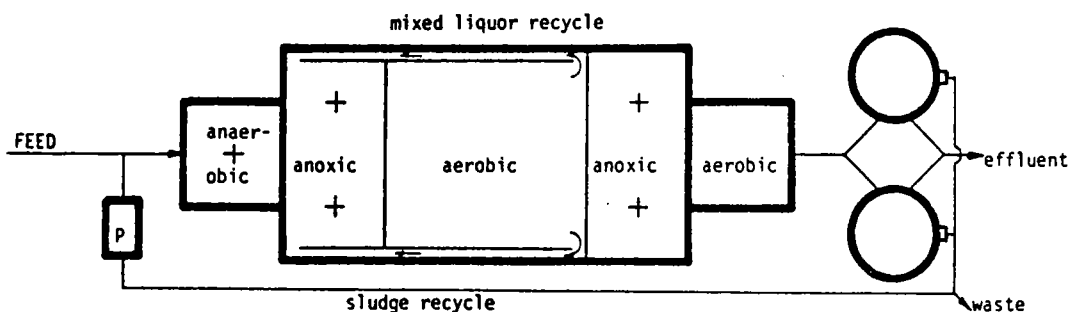

FIG. 8. MODIFIED BARDENPHO PROCESS
(From Barnard, 1976)

To make matters more confusing, in South Africa the Phoredox name is also used to describe the five stage process rather than the anaerobic basin itself. In addition, in some cases the last two basins have been deleted from the five stage process because practical experience has shown that the inclusion of the secondary anoxic basin often results in only slightly greater total nitrogen removals because of the low rate of denitrification under endogenous respiration. The resulting three stage process shown in Fig. 9 is referred to in South Africa as the Modified Phoredox process. Some other processes with the same basic components and different names also exist and will be described later.
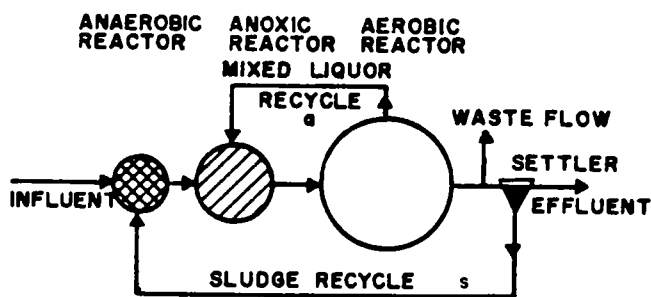

FIG. 9. MODIFIED PHOREDOX PROCESS
(From Ekama et al., 1984)

MECHANISM OF ENHANCED BIOLOGICAL PHOSPHORUS REMOVAL

Some researchers consider most of the phosphorus removal in some of the biological processes described in this paper to be chemical in nature. This will be discussed in greater detail later. However, most researchers agree that under most conditions the major proportion of the phosphorus removal which occurs is biological in nature, i.e. the large amounts of

phosphorus removed are somehow stored in the micro-organisms in the sludge mass. It follows from this that the sink for phosphorus removal in these processes is the waste activated sludge. If activated sludge is not wasted on a regular basis, phosphorus removal will not occur and eventually all phosphorus entering the plant will leave in the effluent.

In all of the processes for enhanced biological phosphorus removal described in this paper, phosphorus release under anaerobic conditions and phosphorus uptake under aerobic conditions are the basic steps. As shown in Fig. 10, if the concentration of soluble orthophosphate in the influent wastewater is between 5 and 10 mg/L as P, the concentration of soluble orthophosphate in solution in the anaerobic zone may be 15 to 30 mg/L as P after release, and around 1.0 mg/L as P in solution in the aerobic zone and final effluent after uptake.
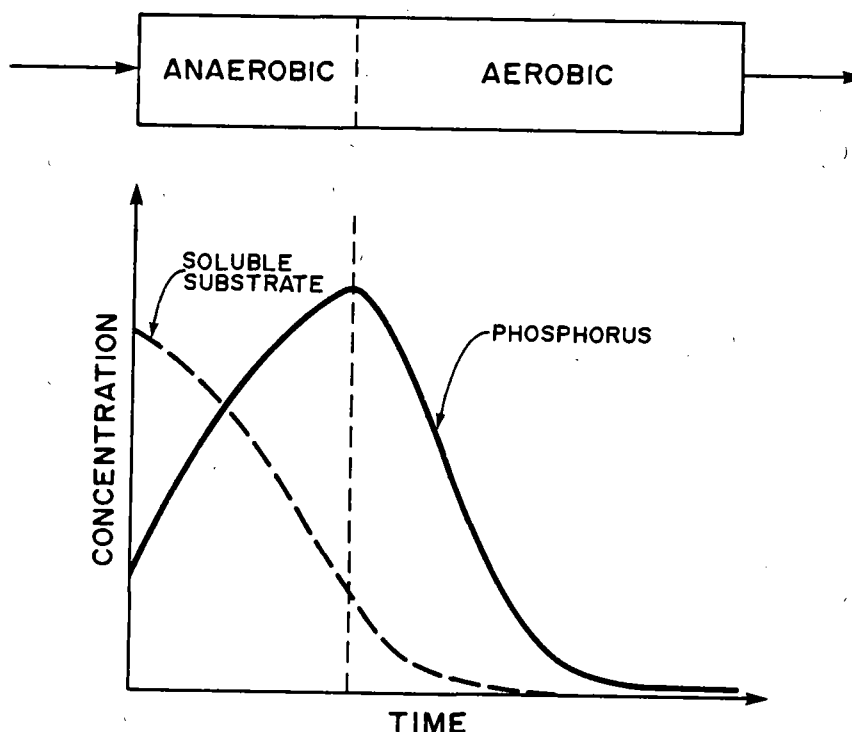
FIG. 10. RELEASE AND UPTAKE OF PHOSPHORUS

For every 100 mg/L of BOD_5 consumed in the process, usually about 1.0 mg/L of phosphorus is required by the micro-organisms for cell growth (Water Pollution Control Federation, 1977). This can be described as normal biological phosphorus removal, and the amount of phosphorus removed in this manner must be subtracted from the overall phosphorus removal to obtain the amount removed by enhanced biological phosphorus removal mechanisms. The uptake of phosphorus in excess of the normal metabolic requirements of the micro-organisms has been referred to in the literature as "luxury uptake", "excess uptake", "enhanced uptake" and "overplus mechanism". Often in the literature it is said that this phenomenon is carried out by the bacteria <u>Acinetobacter</u>, although some researchers have concluded that a number of different bacteria can be involved in enhanced biological phosphorus removal (Osborn and Nicholls, 1978; Florentz and Hartemann, 1984).

General Biochemistry of Biological Phosphorus Removal

There are a number of theories on the precise biochemical mechanisms involved in biological phosphorus removal (see, for example, Comeau (1984) or Evans (1983)). Most of the newer ones state that phosphorus is released by certain micro-organisms under anaerobic conditions in order to provide energy for the rapid intake of readily biodegradable soluble substrates such as fatty acids. Phosphorus which has previously been stored as polyphosphate under aerobic conditions first is transferred to adenosine triphosphate (ATP). In order to produce the energy required to transport soluble substrate across the cell wall into the cell, ATP breaks down to adenosine diphosphate (ADP) and single phosphorus atoms as shown in the following reaction.

$$poly\ P \longrightarrow ATP \longrightarrow ADP + P + energy \qquad (Eq.\ 6)$$
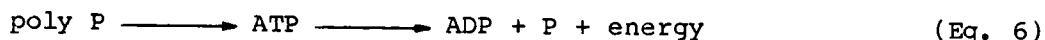
The phosphorus atoms are released from the cell into solution. The soluble substrate taken into the cell under anaerobic conditions is stored as poly-β-hydroxybutyrate (PHB). Under aerobic conditions, this compound is oxidized to provide the energy required for respiration and to drive Equation 6 in reverse as shown below.

$$ADP + P + energy \longrightarrow ATP \longrightarrow poly\ P \qquad (Eq.\ 7)$$
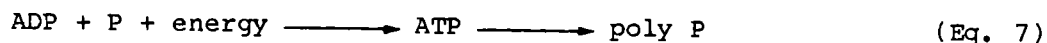
The uptake of phosphorus by micro-organisms in this reaction is the first step in the actual removal of phosphorus, the second being the wasting of the micro-organisms as mentioned previously. The phosphorus taken up is that which has been released into solution under anaerobic conditions, and that entering in the influent wastewater. The amount of phosphorus which can be transported in and out of the micro-organisms by these reactions under alternating anaerobic and aerobic environments is much greater than that entering in the influent wastewater. As a result, very high phosphorus removal efficiencies can be achieved if phosphorus stored in the micro-organisms is regularly withdrawn from the process.

Factors Affecting Biological Phosphorus Removal

It follows from the above discussion that in order for biological phosphorus removal to occur, there must be true anaerobic conditions in the anaerobic basin, i.e. dissolved oxygen and nitrate must be absent, and readily biodegradable soluble substrate must be present. Dissolved oxygen can enter the anaerobic zone in the influent wastewater or return sludge. Nitrate can enter the anaerobic zone in the return sludge if any nitrate produced under aeration is not subsequently removed by denitrification. Nitrate that has entered the anaerobic zone can be removed there by denitrification. However, it is likely that the same soluble substrate required for biological phosphorus removal will be used as the carbon source for denitrification, thereby creating another problem.

Siebritz et al. (1983) have stated that the concentration of readily biodegradable soluble substrate in the anaerobic basin should be greater than 25 mg/L as COD for enhanced biological phosphorus removal to occur. Three methods for measuring the concentration of readily biodegradable soluble substrate have been presented by Nicholls et al. (1984).

DESIGN AND OPERATION OF THE MODIFIED BARDENPHO PROCESS

Typical detention times for the five stages of the Modified Bardenpho process are given in Table 1 (Paepcke, 1983). These are nominal detention times based on the influent wastewater flowrate only.

TABLE 1. TYPICAL DETENTION TIMES FOR MODIFIED BARDENPHO PROCESS

Stage	Detention Time (hr)
Anaerobic	1.3
Primary anoxic	3.2
Aeration	11.2
Secondary anoxic	3.3
Reaeration	1.1
Total	20.1

Mixed liquor is usually recycled from the aeration basin to the primary anoxic basin at a rate of four times the influent flowrate. Return sludge flowrates can vary from 0.5 to 1.5 times the influent flowrate. Modified Bardenpho plants are essentially extended aeration plants. The relatively long total nominal detention time of about 20 hr shown in Table 1 is typical of extended aeration plants. The SRT's of from 10 to 40 d utilized in the Modified Bardenpho process to maintain nitrification are similar to those found in extended aeration plants.

Sludge Treatment and Disposal

Treatment and disposal of the phosphorus rich waste mixed liquor or secondary sludge from a Modified Bardenpho plant or any other biological phosphorus removal process requires special consideration. This is because phosphorus may be released from the sludge into the environment if the sludge is exposed to anaerobic conditions. Generally then, the sludge should be kept aerobic during handling and treatment, and the site for final disposal of the sludge should be at a location where any liquid released from the sludge will not be directly discharged to a watercourse.

Thickening. Dissolved air flotation has been used for the thickening of waste mixed liquor from biological phosphorus removal plants because it is a process in which the biological solids are maintained in an aerobic environment. Gravity thickening in a simple circular clarifier-like device has been employed at a number of plants in South Africa. If the supernatant from this treatment step contains only a minimal amount of phosphorus, it can be returned to the plant influent. Otherwise, it could be applied directly to agricultural land or undergo chemical precipitation.

Dewatering. Vacuum filters, belt filter presses or plate and frame presses may be used for dewatering of sludges from biological phosphorus removal plants. However, any sidestreams resulting from the use of these processes must be handled appropriately if they contain significant amounts of phosphorus.

Anaerobic Digestion. In cases where existing treatment plants with anaerobic digestion are being retrofitted for biological phosphorus removal, it may be deemed desirable from an energy point of view to continue anaerobic digestion of waste sludges. The supernatant from anaerobic digesters treating waste mixed liquors or secondary sludges rich in phosphorus will likely contain a considerable amount of phosphorus, and this should be removed by chemical precipitation prior to the return of the supernatant to the plant.

APPLICATION OF THE MODIFIED BARDENPHO PROCESS IN SOUTH AFRICA

Subsequent to Barnard's pilot-scale experiences in which he observed excellent nitrogen and phosphorus removal without the addition of chemicals (Barnard, 1974), the City of Johannesburg, South Africa conducted experiments at two of its full-scale extended aeration plants to see if similar results could be achieved. At the Alexandra Works shown in Fig. 11, denitrification of 98 percent of the nitrate formed and an overall nitrogen removal of 85 percent were achieved by turning off aerators at the inlet end of the plant near where the influent wastewater and return sludge entered in order to create an anoxic zone (Nicholls, 1975). Nitrified mixed liquor was recycled from the aerobic portion of the first basin to this anoxic zone by the mixing action of the aerators which were still running. Turning off all the aerators for five hours during the night created anaerobic conditions and resulted in an average overall phosphorus removal of 46 percent.
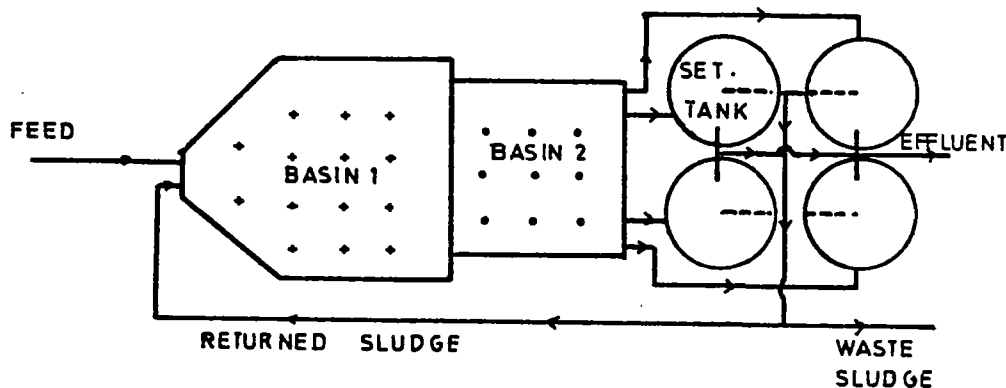

FIG. 11. ALEXANDRA WORKS, JOHANNESBURG
(From Nicholls, 1975)

Extensive experimentation was carried out at the Olifantsvlei Works shown in Fig. 12 (Venter et al., 1978). Various combinations of aerators in the first three compartments were turned off in order to create anoxic and anaerobic zones. Soluble orthophosphate concentrations as low as 0.9 mg/L as P were observed in the final effluent. Additional advantages of this mode of operation included reduced power costs and elimination of rising sludge due to denitrification in the secondary clarifiers.

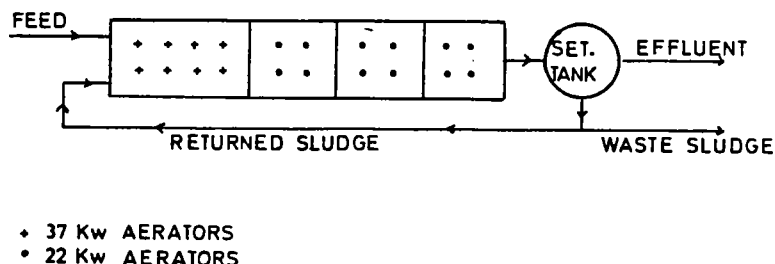* 37 Kw AERATORS
* 22 Kw AERATORS

FIG. 12. OLIFANTSVLEI WORKS, JOHANNESBURG
(From Nicholls, 1975)

The results of the experimentation at the Alexandra and Olifantsvlei Works led to the decision to employ the five stage Modified Bardenpho process in the design of two new large treatment plants in Johannesburg known as the Goudkoppies and Northern Works. Both plants had three parallel modules with design capacities of 50 000 m^3/d per module, and became operational in 1978 and 1980 respectively. During the late 1970's, a number of smaller but similar treatment plants were constructed at other locations around South Africa. These plants were built in anticipation of a limit on soluble orthophosphate of not greater than 1.0 mg/L as P which came into effect in August 1980 for certain areas of South Africa. Prior to the construction of these new facilities designed for biological nitrogen and phosphorus removal, there were few activated sludge plants in South Africa. Most of the municipal wastewater treatment plants were trickling filters. By 1980 there were about fifteen full-scale municipal treatment plants in South Africa whose designs were based on the Modified Bardenpho process. In addition, there were another ten facilities which were operated for biological nitrogen and/or phosphorus removal, including the Alexandra and Olifantsvlei Works of Johannesburg discussed above. There were also about four oxidation ditches whose aeration patterns were altered to achieve biological nutrient removal. Also, there were one or two sites at which phosphorus removal by chemical addition was practised. It is estimated that since that time up to ten additional full-scale multi-stage activated sludge plants designed for biological nitrogen and phosphorus removal have been built in South Africa.

PERFORMANCE AND OPERATIONAL ASPECTS OF BIOLOGICAL PHOSPHORUS REMOVAL PLANTS IN SOUTH AFRICA

In 1980 and 1981, the author carried out a survey of eleven full-scale treatment plants in South Africa whose designs were based on the Modified Bardenpho process (Paepcke, 1983). These plants represented the majority

of the full-scale plants in South Africa which were designed for biological nitrogen and phosphorus removal and in operation at the time. The objectives of the survey were to identify those plants which successfully removed nitrogen and phosphorus and to establish the operating problems of those that did not. The results of the survey, particularly with respect to phosphorus removal, are summarized below.

Technical Description of Plants

Plant Configurations. A typical plant layout representative of all eleven plants covered in the survey is shown in Fig. 13. The configurations of some of the plants in the survey, however, were not exactly as shown in Fig. 13. Two of the plants did not have a secondary anoxic basin. In one of these, the primary anoxic and aeration zones were not physically separated by a wall. In one of the plants the five stages were placed in an annular ring around the secondary clarifier. One of the plants was divided into parallel halves which had slightly different aeration and secondary anoxic basin volumes. The two largest plants both consisted of three separate parallel five-stage modules.
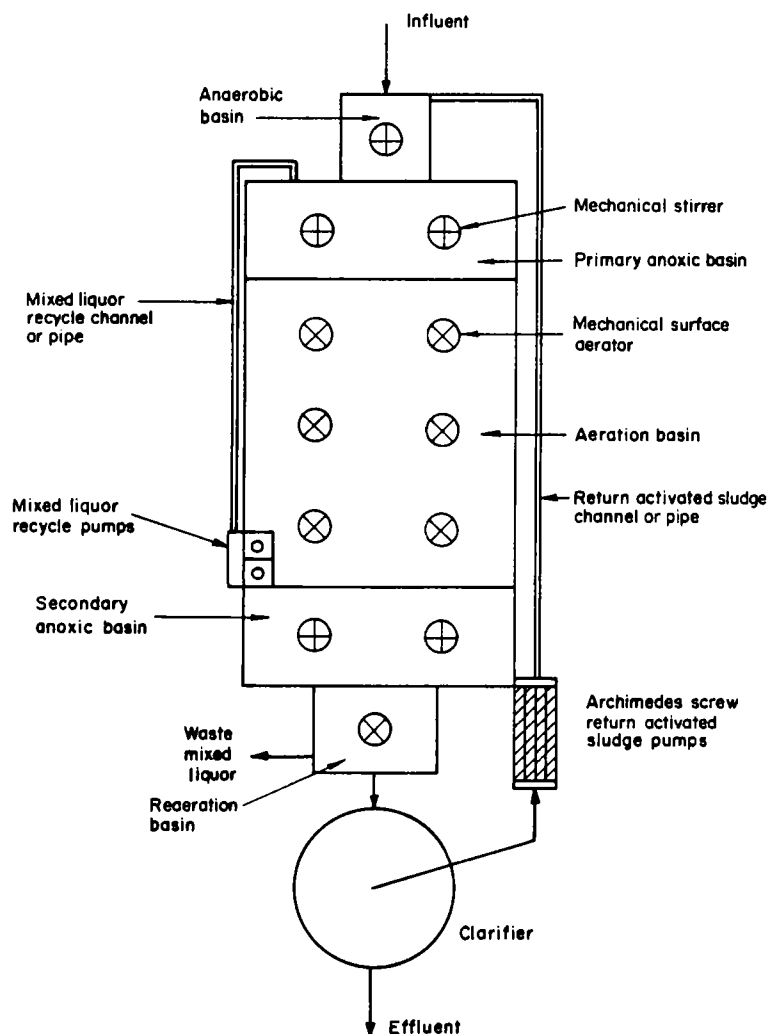

FIG. 13. TYPICAL PLANT LAYOUT
(From Paepcke, 1983)

<u>Preliminary and Primary Treatment.</u> In each case the influent wastewater underwent screening, grit removal and flow measurement before entering the plant. The three largest plants had primary settling basins, and two of these also had flow equalization tanks.

<u>Nominal Detention Times.</u> The minimum, maximum and mean values for the nominal detention times, based on the design flow rate, for the five stages of the eleven plants in the survey are given in Table 2.

TABLE 2. <u>NOMINAL DETENTION TIMES</u>

Basin	Min.(hr)	Max.(hr)	Mean(hr)
Anaerobic	0.6	1.9	1.3
Primary anoxic	2.2	5.2	3.2
Aeration	6.7	19.0	11.2
Secondary anoxic	2.2	5.7	3.3
Reaeration	0.5	1.6	1.1
Total	14.0	30.3	20.1

<u>Aeration and Stirring.</u> In ten of the eleven plants the aeration was by mechanical surface aerators. Only one of the plants used a diffused air system in the aeration basin. Diffused air was used for reaeration in four of the eleven plants, and mechanical surface aerators were used in the others. Mechanical stirrers were used to mix the contents of the anaerobic and anoxic basins in ten of the plants. In the other plant, which was the one utilizing diffused air for aeration, the stirring action was caused by the flow of return activated sludge or recycled mixed liquor being pumped into the anaerobic and anoxic basins through nozzles.

<u>Mixed Liquor Recycle.</u> In seven of the plants, mixed liquor was recycled from the aeration basin to the primary anoxic basin through a channel or pipe by pumps with capacities varying from 3.5 to 16 times the design flow rate of the plant. In the other plants the recycling of mixed liquor from the aeration basin to the primary anoxic basin was through openings in the wall between the two basins by the pumping action of aerators in the aeration basin.

<u>Return Activated Sludge Pumping.</u> In nine of the eleven plants, Archimedes screw type pumps were used for the pumping of return activated sludge from the final clarifiers to the anaerobic basin. In the other two plants other types of pumps were used. The return sludge pumping rate was generally 1.0 or 1.5 times the design flowrate.

<u>Solids Retention Time.</u> The solids retention time (SRT) in the plants surveyed was commonly controlled by withdrawing a constant volume of waste mixed liquor daily from either the aeration or reaeration basin. Solids retention times of from 20 to 30 d were used.

Dissolved Oxygen Control. All the plants had some means to control the dissolved oxygen concentration in the aeration basin. Six of the plants had a weir of adjustable height which controlled the level of the mixed liquor in the aeration basin. A higher level of mixed liquor resulted in a greater depth of immersion of the surface aerators and thus a greater degree of aeration. The weir height was either set manually or adjusted by an automatic control system connected to dissolved oxygen measuring probes in the aeration basin. In four of the plants, surface aerators were turned on and off in order to control the dissolved oxygen concentration in the aeration basin. Some had timing mechanisms which were manually preset to turn aerators on and off at fixed intervals, while others had automatic systems incorporating dissolved oxygen measuring probes. The plant which utilized diffused air had an automatic control system regulating the output of the blowers in order to maintain the dissolved oxygen concentration within certain limits.

Results of Survey

Information on the operation and performance of the eleven plants in the survey is summarized in Table 3. The plants are numbered in the chronological order in which they were visited. The information in Table 3 was up to date as of August 1981.

The amount of influent and effluent quality parameter data available varied from plant to plant. Generally the analytical records for the year prior to the visit to the plant were inspected at the time of the visit. In some cases a series of regular sample analyses was available for a shorter period only. The influent and effluent quality parameter values shown in Table 3 are average values considered to have been representative of the operation of the plant for a significant period. Those cases where only a minimal amount of data were available are noted in Table 3.

Design and Actual Flowrates. Of the eleven plants in the survey, seven had design flowrates of less than 10 000 m^3/d, two had design flowrates between 10 000 and 20 000 m^3/d, and there were two large plants each with a total capacity of 150 000 m^3/d. On average the plants in the survey had been operating for only two or three years, and were treating about 60 percent of their design flow capacities. Only two of the eleven plants were overloaded. Because none of the plants was operating at its design flowrate, it was difficult to evaluate aspects of the designs related to flowrate such as nominal detention times, and it was impossible to have known what the performances of the plants would have been at full hydraulic load. It should also be noted that in no case was it evident that plant underloading was a factor responsible for poor performance.

Wastewater Origins. The influent to plants 5, 7 and 11 contained a significant proportion of industrial wastewater. The other eight plants treated domestic wastewater.

COD Removal. The average chemical oxygen demand (COD) of the influent wastewater at the eleven plants ranged from 400 to 500 mg/L although this, of course, varied somewhat from site to site. Inspection of Table 3 reveals that, with the exception of two cases, all of the plants produced an effluent with a COD concentration less than the South African General Standard of 75 mg/L. This indicates that COD removal was successful.

TABLE 3. RESULTS OF SURVEY
(From Paepcke, 1983)

Plant no.		1	2	3
Configuration		Aerobic and anoxic zones not physically separate	Five-stage	Five-stage
Date of commissioning		September 1976	March 1978	October 1979
Design flow		3.5 Mℓ/d	7.0 Mℓ/d	9.0 Mℓ/d
% Loading		140	60	Varies - underloaded
Effluent (mg/ℓ)	COD	50	35	60 to 100
	NH₃-N	2.0	0.4	3.0
	NO₃-N	<1.0	4.0	Varies up to 15
P (mg/ℓ as P)	Infl.	3.0 to 10 (ortho P)	13 (total P)	2.0 to 7.0 (ortho P)
	Effl.	1.0 (ortho P)	9.0 (total P)	2.0 to 7.0 (ortho P)
COD:TKN ratio		15:1	8.3:1	COD = 400 NH₃ = 20 to 30
Aeration points			Archimedes screws cascades, vortices	
Mechanical problems				Mixed liquor recycle pumps, severe corrosion in clarifier, clarifier scraper wheel
R E M A R K S		- values given are for performance over past 9 months - occasionally N and P removal not as good as shown, deteriorates and then takes time to come back - in the previous year no P removal and inconsistent nitrification - improved performance thought to be due to better control of waste sludge quantities (SRT) in past 9 months	- poor P removal due to high DO in anaerobic basin - this due to air entrainment at the three aeration points and in addition the influent sewage has a high DO because of steep gradient in collection system	- stable operation of plant not yet achieved due to mechanical problems

Effluent (mg/ℓ) — NH_3-N, NO_3-N; P (mg/ℓ as P); COD:TKN ratio with NH_3 = 20 to 30

TABLE 3.   RESULTS OF SURVEY (CONTINUED)
(From Paepcke, 1983)

Plant no.		4	5	6
Configuration		Five-stage	Five stages in annular ring around clarifier	Five-stage Two parallel halves
Date of commissioning		April 1977	June 1979	August 1979
Design flow		5.0 Mℓ/d	4.5 Mℓ/d	7.8 Mℓ/d
% Loading		60	60 to 75	60
Effluent (mg/ℓ)	COD	Varies below 75	90	40
	NH₃-N	10	<1.0	0.9
	NO₃-N	Usually below 1.5	<1.5	6.0
P (mg/ℓ as P)	Infl.	9.0 (total P)	15 (ortho P)	10 (total P)
	Effl.	1.0 to 7.0 (ortho P)	6.0 to 10 (ortho P)	7.0 (total P)
COD:TKN ratio		COD = 500 NH₃ = 30	COD = 880 NH₃ = 20	COD = 350 NH₃ = 40
Aeration points				Archimedes screws cascades
Mechanical problems		Many in past – aerators, pumps, etc.	Breakages in aeration piping and diffusers, blowers, DO control system	Many in past – aerators, stirrers, Archimedes screws
R E M A R K S		– in recent months N and P removal has been much better than values given – this is due to better DO control following the installation of an automatic timing system for the aerators	– parameter values given are based on a limited number of samples – several grab samples in recent months indicate good phosphorus removal to about 1.0 mg/ℓ in effluent – reason for prior poor performance is not clear	– incomplete denitrification because of low COD:TKN ratio ($\leq$ 7:1) in sewage – poor phosphorus removal because of DO and NO₃ in anaerobic basin – DO in anaerobic basin because of air entrainment at aeration points – NO₃ introduced into anaerobic basin in return sludge

TABLE 3.   RESULTS OF SURVEY (CONTINUED)
(From Paepcke, 1983)

Plant no.		7	8	9
Configuration		Four-stage no secondary anoxic	Five-stage	Five-stage
Date of commissioning		February 1979	March 1979	August 1979
Design flow		8.0 Mℓ/d	10.5 Mℓ/d	16.0 Mℓ/d
% Loading		200 incl. filter effluent	50	60
Effluent (mg/ℓ)	COD	50	-	50
	NH_3-N	20	-	1.0
	NO_3-N	4.0	-	14
P (mg/ℓ as P)	Infl.	unknown	-	15 (ortho P)
	Effl.	1.0	-	11 (ortho P)
COD:TKN ratio		unknown	-	6.6:1 after primary settling
Aeration points				Archimedes screws cascades, vortices
Mechanical problems			Minor	Minor
R E M A R K S		- parameter values given are based on a limited number of grab samples - plant receives effluent from adjacent biological filtration works which increases hydraulic and nitrogen load	- not enough sample analyses available to assess performance of plant	- incomplete denitrification because of low COD:TKN ratio in settled sewage - poor phosphorus removal because of DO and NO_3 in anaerobic basin - DO in anaerobic basin because of air entrainment at aeration points - NO_3 introduced into anaerobic basin in return sludge

TABLE 3. <u>RESULTS OF SURVEY (CONTINUED)</u>
(From Paepcke, 1983)

Plant no.		10	11
Configuration		Five-stage Three modules	Five-stage Three modules
Date of commissioning		January – May 1980	May 1978
Design flow		150 Ml/d total	150 Ml/d total
% Loading		50	Modules 1 & 2 – 78 Module 3 – 42
Effluent (mg/l)	COD	60	60
	NH$_3$-N	1.5	4.0
	NO$_3$-N	8.8	4.0
P (mg/l as P)	Infl.	9.4 (total P)	7.5 (total P)
	Effl.	7.1 (total P)	1.0 (ortho P)
COD:TKN ratio		10.3:1 after primary settling	12.1:1 after primary settling
Aeration points			Archimedes screws cascades, vortices
Mechanical problems			Many in the beginning
R E M A R K S		- incomplete denitrification because of low COD:TKN ratio in settled sewage - poor phosphorus removal because NO$_3$ introduced into anaerobic basin in return sludge	- because of low COD:TKN ratio in settled sewage have to limit nitrification to about 4.0 mg/l NH$_3$ remaining in order to minimize effluent NO$_3$ - plant subjected to a high degree of control

Nitrification. The wastewater to be treated at the eleven plants generally
had a total Kjeldahl nitrogen (TKN) concentration ranging from 40 to 60
mg/L as N and an ammonia concentration from 20 to 40 mg/L as N.
Inspection of Table 3 reveals that, with the exception of one case, the
ammonia concentration in the plant effluents was less than or equal to the
South African General Standard of 10 mg/L as N.

Denitrification. Denitrification in the eleven plants was not as
consistent as the COD removal and nitrification as evidenced by the
effluent nitrate concentrations given in Table 3. The nitrate
concentrations varied from less than 1.0 mg/L as N to 15 mg/L as N. Plant
2 was the only plant in which both nitrification and denitrification had
been proved to be consistently good.

Phosphorus Removal. As shown in Table 3, the total phosphorus
concentration in the influent wastewaters ranged from 7.5 to 13 mg/L as P,
and the orthophosphate concentration (in those cases where orthophosphate
was measured) varied from 2.0 to 15 mg/L as P. The last plant in Table 3
was the only one of the eleven to have proved its capability to remove
phosphorus consistently to a level of 1.0 mg/L orthophosphate as P or less,
which was the effluent standard in effect in some areas of South Africa.
    However, it should be noted that in the few months before the survey
was finished, phosphorus removal in plants 1, 4 and 5 improved vastly to
about 1.0 mg/L orthophosphate as P in the effluent, primarily due to better
control procedures. Thus, while these plants had not yet proved their
capability to remove phosphorus consistently to this low level, their
recent performance did indicate good potential.

Factors Affecting Phosphorus Removal
    The major factors which adversely affected phosphorus removal in the
plants surveyed were excessive amounts of nitrate and free dissolved oxygen
in the anaerobic basins. The presence of nitrate or free dissolved oxygen
in the anaerobic basin prevented the formation of an anaerobic condition
and the release of phosphorus which is necessary for subsequent phosphorus
removal in the aeration basin.

Presence of Nitrate in Anaerobic Basin. In plants 3, 6, 9 and 10 which
experienced incomplete denitrification, nitrate was introduced into the
anaerobic basin in the return activated sludge. In these plants virtually
complete denitrification had to occur before phosphorus removal could be
successful.
    At plants 6, 9 and 10 the major factor responsible for poor
denitrification was thought to be a low COD:TKN ratio in the wastewater to
be treated. It was generally considered that the COD:TKN ratio must be at
least 10:1 for denitrification to be complete in this type of design. This
ratio ensured that there was adequate carbon and energy available to
maintain the level of activity of the denitrifying micro-organisms in the
primary anoxic basin required to denitrify all the nitrate produced if
nitrification was complete. A low COD:TKN ratio was also an important
factor in the operation of plant 11. In order to minimize the nitrate
concentration in the plant effluent and return activated sludge,

nitrification was limited to about 4.0 mg/L NH_4^+ remaining in the plant effluent by restricting the oxygen input. Otherwise there would have been too much nitrate for complete denitrification at the low COD:TKN ratio.

At plants 9, 10 and 11, the three largest, the raw wastewater underwent primary settling before entering the activated sludge units. In these cases the raw wastewater had a COD:TKN ratio favourable for denitrification. Unfortunately, during primary settling the COD:TKN ratio was lowered from that in the raw wastewater as more carbonaceous matter was removed than nitrogenous matter. This was because the carbonaceous matter in the raw wastewater was predominantly in the particulate form while the nitrogenous matter was mainly soluble. At plant 9, for a trial period, the primary settling basins were bypassed and the plant was fed with raw wastewater. This resulted in an improvement in phosphorus removal. At plant 10, it was not possible to bypass the primary settling basins because of the physical layout of the works. Instead, however, for a trial period the feed to the plant was supplemented with sludge from the primary settling basins in order to increase the COD. This was also successful. The same procedure was to be tried at plant 6 using raw primary sludge from an adjacent trickling filter plant. At plant 11 it was planned to construct a cross-connection between two sewers so that the wastewater entering the plant had a greater industrial component and a higher COD.

The COD:TKN ratios in the settled wastewater at plants 10 and 11 averaged 10.3:1 and 12.1:1, respectively. It was interesting to note that these values were considered to be unfavourably low for complete denitrification to occur in these two plants, although they were greater than the figure of 10:1 generally considered necessary. This inconsistency is possibly an indication that the COD:TKN ratio was not the best parameter for assessing the suitability of wastewater for denitrification in this type of design. Perhaps it would have been better to measure that fraction of the carbonaceous matter in the influent wastewater that was soluble and readily available. However, this is not easily done, and the importance of this type of substrate, particularly with respect to phosphorus removal, was not as widely recognized at that time as it is now.

The denitrification in plant 3 was poor because the mixed liquor recycle pumps were out of service for a considerable part of the time due to a number of mechanical and electrical problems.

Presence of Dissolved Oxygen in Anaerobic Basin. A number of reasons were identified for the presence of excess free dissolved oxygen in the anaerobic basins of plants 2, 6, 9 and 11. The dissolved oxygen concentration in the return activated sludge was observed to increase by as much as 5 mg/L as it passed through the Archimedes screw return activated sludge pumps. This was because as the return sludge passed over the surface of the screws, it was continually exposed to the atmosphere. This dissolved oxygen was then introduced into the anaerobic basin in the return activated sludge. In addition, in these four plants there were cascades in the influent or return sludge channels, or both, which created turbulence and thus caused entrainment of oxygen which was then introduced into the anaerobic basin. In several cases vortices created by the stirrers were also suspected of entraining air into the basins.

The entrainment of oxygen during return sludge pumping could probably have been avoided by using a different type of return activated sludge pump, or if Archimedes screws were used, by placing them as far away from

the anaerobic basin as possible so as to allow time for the entrained oxygen to be consumed by the normal respiration of the sludge. This problem might also have been overcome by enclosing the Archimedes screws in an oxygen- free environment. Cascades should have been avoided in the design of the plants. In the past a stirring energy input of about 10 W/m^3 had been used, but it is now thought that as little as 2.5 W/m^3 is necessary. The use of lower stirring energy would eliminate vortices and air entrainment from this source. In addition, the introduction of both dissolved oxygen and nitrate into the anaerobic basins would have been a less significant factor if longer anaerobic detention times had been used.

Other Factors Affecting Phosphorus Removal. A large number of various mechanical problems in plants 3, 4, 5 and 6 interrupted stable operation and affected plant performance for significant periods. These problems are, to a certain degree, unavoidable during the early operation of a new wastewater treatment plant. Thus these plants had not been in operation long enough for their phosphorus removal capability to be fully evaluated.

Operator training is always important at a wastewater treatment plant. Proper operation of the plants in the survey probably required a higher standard of training and a greater theoretical knowledge of the process on the part of the operators than that required for the successful operation of a simpler plant such as a trickling filter or conventional activated sludge plant. The performance of some of these eleven plants undoubtedly would have been better if the operators had had a higher standard of training and a greater theoretical knowledge of the process.

In the case of plants 5, 7 and 8, and to a certain extent in other cases also, there was a scarcity of routine sample analyses. This made a proper assessment of the performances of the plants difficult, and obviously prevented the development of rational operating policies.

Because the biological processes for nitrogen and phosphorus removal are more complex than other biological wastewater treatment processes, they are generally more sensitive to factors such as changes in wastewater characteristics, minor mechanical breakdowns or lack of operator attention and thus are more prone to upsets. This in general affected nitrogen and phosphorus removal in the plants in the survey.

In the light of the three preceding points, it should be noted that plant 11, the only one of the eleven to have consistently removed phosphorus, was situated at a large municipality which had a relatively large and highly skilled staff to oversee its operation. The plant was subjected to an extensive degree of control and a comprehensive monitoring program was carried out regularly. This was necessary for continued successful operation because of the sensitivity of the process. Thus it might have been difficult to achieve this kind of success at a small municipality having a minimal staff with limited training and a lesser facility for carrying out sample analyses.

Conclusions

Based on the information generated during the survey, the following conclusions were made.

1. Only one of the eleven plants surveyed had proved its capability for consistent phosphorus removal to a low level although three others had achieved very good phosphorus removal in recent months.

2. The major factors adversely affecting phosphorus removal in the plants covered by the survey were excessive amounts of nitrate and dissolved oxygen in the anaerobic basins. Nitrate was introduced into the anaerobic basin in the return activated sludge when denitrification was incomplete. The major factor responsible for incomplete denitrification was a disproportionately low amount of carbonaceous matter in the raw or settled wastewater relative to the total nitrogen concentration. The dissolved oxygen was the result of air entrainment at the Archimedes screw return activated sludge pumps, at cascades in the influent and return sludge channels, and by vortices caused by the stirrers.

3. Other factors which significantly affected phosphorus removal were mechanical problems, inadequate operator training and insufficient sample analyses on which to base operating decisions.

4. The relatively highly skilled staff and large number of sample analyses possibly required for successful operation of a biological phosphorus removal plant of this type could well be a major burden for some small municipalities in South Africa.

It should be noted that these conclusions were based on South African experience which may or may not be directly applicable to other countries including Canada. Domestic wastewaters in Canada are not as strong as those in South Africa and other conditions may be significantly different. It should also be emphasized that the information upon which these conclusions were based is up to date only as of August 1981. Unfortunately, a comprehensive data base on the operation of these plants since then and those newer plants which have become operational since 1981 is not available. It would be reasonable to assume that all of the experience gained prior to 1981 would result in improved operation and performance thereafter, and that this experience was also incorporated in the design of the newer plants built since then. However, an evaluation of the South African experience prior to 1981 is valuable to Canadians who may be contemplating the construction of a biological phosphorus removal plant as it reveals some of the problems that can occur if all precautions are not taken.

Recent Developments in South Africa
One of the new approaches that was considered as a result of these early experiences with biological phosphorus removal at full-scale in South Africa was augmenting the influent wastewater with an external source of the readily biodegradable soluble substrate whose presence in the anaerobic zone is deemed necessary for enhanced biological phosphorus removal to occur. Experiments were carried out at the Northern Works in Johannesburg in which the supernatant from acid digestion of primary sludge was added to the anaerobic zone (Nicholls et al., 1984). This had been suggested as early as 1977 (Osborn and Nicholls, 1978). The supernatant was found to contain significant quantities of volatile fatty acids which are a readily biodegradable soluble substrate. The mass of phosphorus in the digester

supernatant effectively increased the total phosphorus concentration in the influent wastewater from 11.0 to 17.5 mg/L, and an effluent soluble orthophosphate concentration of 2.2 mg/L was obtained. This represented a phosphorus removal of about 15 mg/L which was a significant improvement over previous performance at the plant. The addition of fermentation products from the acid digestion of primary sludge also plays an important role in the operation of the Kelowna Pollution Control Centre as will be discussed elsewhere at this seminar.

Since the survey was conducted, another large treatment plant designed for biological nitrogen and phosphorus removal has been constructed in Johannesburg. It is known as the Bushkoppie Works and uses the three stage process known as Modified Phoredox (Fig. 14). The anaerobic basin is divided into two compartments and the aeration basin is "folded" in order to reduce the capital cost of the mixed liquor recycle channel. Unlike the other large plants in South Africa, aeration is by diffused air and volatile fatty acids can be produced by acid digestion on a regular basis. Operating results are not yet available.
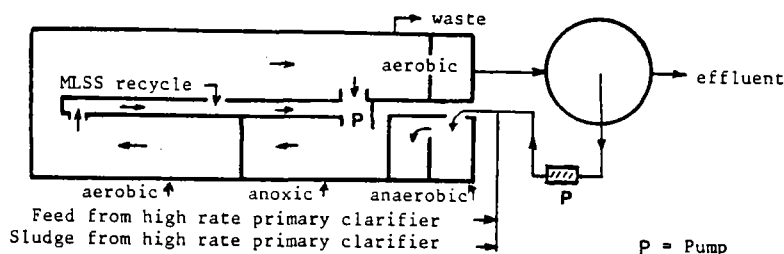

FIG. 14. BUSHKOPPIE WORKS, JOHANNESBURG
(From Osborn et al., 1979)

APPLICATION OF THE MODIFIED BARDENPHO PROCESS IN NORTH AMERICA

All of the known full-scale Modified Bardenpho plants in the world outside of South Africa are located in North America. The Kelowna Pollution Control Centre is the only one in Canada. There are four Modified Bardenpho plants in operation in the United States. These are located in Palmetto, Florida; Hills Development, New Jersey; Payson, Arizona and Orange County, Florida. There are two under construction in Fort Myers, Florida and another one under construction in Tarpon Springs, Florida (Di Gregorio, 1985).

Palmetto Modified Bardenpho Plant

The first Modified Bardenpho plant in North America began operation in Palmetto, Florida in October 1979. It has a design capacity of 5300 m^3/d and is divided into two parallel modules as shown in Fig. 15. The nominal detention times for the five zones of this plant are given in Table 4.
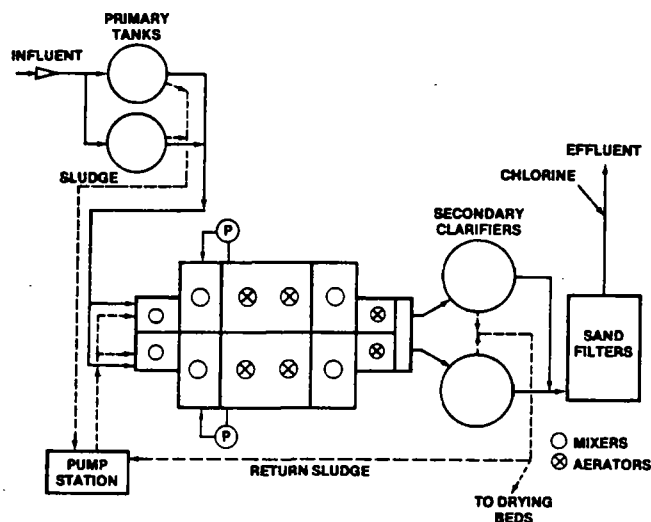
FIG. 15. PALMETTO MODIFIED BARDENPHO PLANT
(From Burdick et al., 1982)

TABLE 4. DETENTION TIMES FOR PALMETTO MODIFIED BARDENPHO PLANT

Basin	Detention Time (hr) (Based on $Q = 5300$ m^3/d)
Anaerobic	1.0
Primary anoxic	2.7
Aeration	4.7
Secondary anoxic	2.2
Reaeration	1.0
Total	11.6

Phosphorus removal data from Burdick et al. (1982) and Irvine et al. (1982) for three different periods of operation are shown in Table 5. During the third period from October 1981 to March 1982, a minimal alum dosage was added to the secondary clarifiers to ensure that the effluent total phosphorus concentration was 1.0 mg/L or less. The influent BOD_5 concentration averaged 135 mg/L. In terms of COD, this would be about 270 mg/L which is much lower than that at South African plants. Lower effluent phosphorus concentrations probably would have been achieved at Palmetto if the readily biodegradable soluble substrate concentration in the influent wastewater was higher.

TABLE 5. PHOSPHORUS REMOVAL AT PALMETTO, FLORIDA

Total Phosphorus (mg/L)	Jan. 1980 to April 1980	April 1981 to Sept. 1981	Oct. 1981 to March 1982
Raw wastewater	8.2	6.2	7.8
Secondary clarifier effluent	2.9	--	--
Sand filter effluent	2.9	2.2	1.0
% Removal	65	65	87*

* Minimal alum dosage applied

UNIVERSITY OF CAPE TOWN (UCT) PROCESS

A process developed at the University of Cape Town in South Africa and known as the UCT process is shown in Fig. 16. It is similar to the three stage Modified Phoredox process except for that the return sludge is discharged to the anoxic reactor rather than the anaerobic reactor, and mixed liquor is recycled from the anoxic reactor to the anaerobic reactor. Its principal advantage over the Modified Phoredox and Modified Bardenpho processes is that there is an opportunity for any nitrate in the return sludge to be denitrified in the anoxic basin prior to entry to the anaerobic basin. It is claimed (Ekama et al., 1984) that enhanced phosphorus removal in the Modified Bardenpho process will deteriorate at an influent TKN/COD ratio greater than 0.08 (or a COD/TKN ratio less than 12:1), while phosphorus removal in the UCT process can be maintained at a TKN/COD ratio up to about 0.13 (or a COD/TKN ratio as low as 7.5:1). It should be noted that these figures apply to typical South African wastewaters, and are not relevant if the wastewater is modified by addition of supernatant containing readily biodegradable soluble substrate. In contrast to these observations, Gerber et al. (1982) found in pilot-scale studies that at a COD:TKN ratio of 9.5:1 the UCT process had no advantage over the Modified Bardenpho configuration.
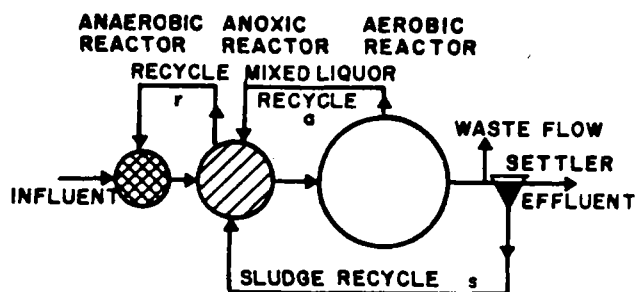

FIG. 16. UCT PROCESS
(From Ekama et al., 1984)

There is also a Modified UCT process as shown in Fig. 17. In this configuration the anoxic reactor is divided into two separate compartments so that denitrification of the return sludge and mixed liquor recycle streams can be controlled separately. At a TKN/COD ratio above 0.11 (or a COD/TKN ratio below 9:1), nitrate begins to appear in the mixed liquor recycled to the anaerobic reactor in the Modified UCT process.
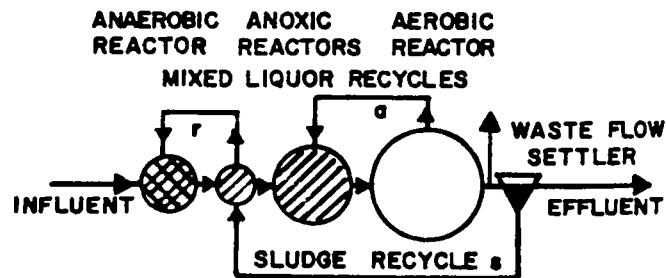

FIG. 17. MODIFIED UCT PROCESS
(From Ekama et al., 1984)

There is also a process combination known as the Modified UCT/UCT process as shown in Fig. 18. It features an optional recycle of mixed liquor from the second anoxic reactor to the anaerobic reactor. This allows the configuration to be operated as either the UCT process or Modified UCT process.
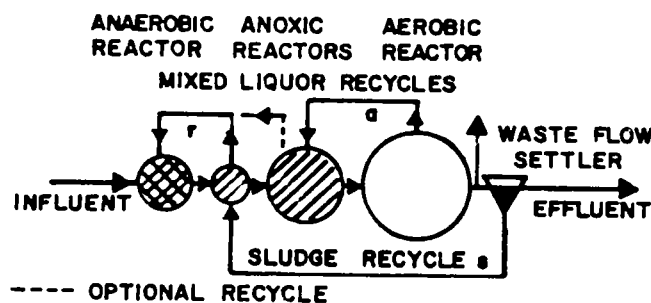

FIG. 18. MODIFIED UCT/UCT PROCESS
(From Ekama et al., 1984)

The UCT process has been tested extensively at laboratory and pilot-scale. However, there is only one known full-scale plant in the world. It is in South Africa and is designed so that it can also be

operated as a Modified Phoredox process. It is possible that there are one or two other full-scale plants under construction or in operation in South Africa that are based on the UCT process.

PINETOWN

As mentioned earlier, most reseachers agree that under most conditions the major proportion of enhanced phosphorus removal is biological in nature, i.e. the large amounts of phosphorus removed are somehow stored in the micro-organisms in the sludge mass. An exception is the Umhlatuzana Works in Pinetown, South Africa, where it was claimed (Kerdachi and Roberts, 1983) that phosphorus was removed by extracellular chemical precipitation rather than by intracellular uptake.

This plant was a simple rectangular extended aeration unit with six mechanical surface aerators as shown in Fig. 19. It had a design capacity of 4800 m^3/d and treated raw unsettled wastewater. The wastewater was relatively soft, having an alkalinity of 195 mg/L as $CaCO_3$. The influent total phosphorus concentration averaged 10.5 mg/L and the COD/TKN ratio was 14:1. It was operated with SRT's up to 69 d. Surface aerators were turned on and off with the diurnal variation in influent waste load, and the dissolved oxygen concentration in the basin was always 0.5 mg/L or less.
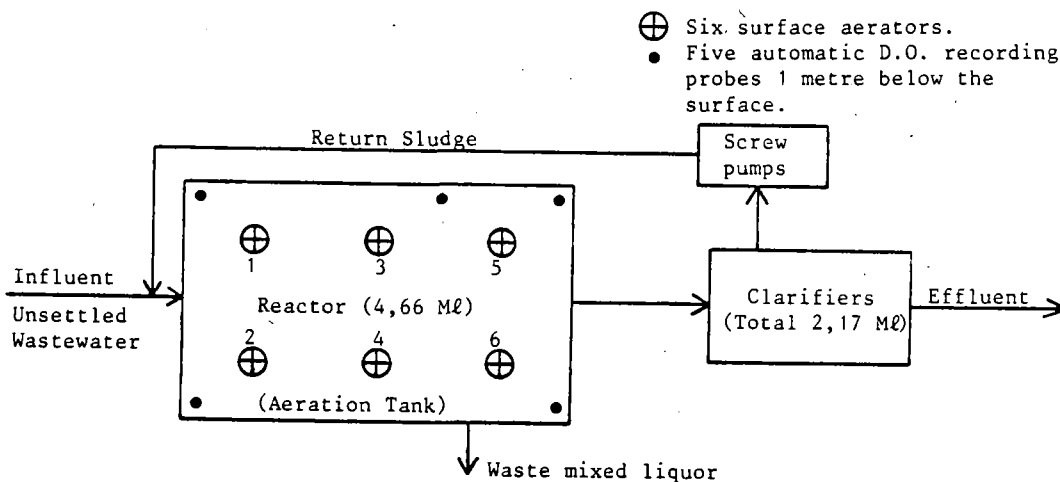
FIG. 19. UMHLATUZANA WORKS, PINETOWN
(From Kerdachi and Roberts, 1983)

Nitrogen and phosphorus removal greater than 90 percent was observed at the longer SRT's. It was postulated that nitrification and denitrification took place in the same basin in numerous pockets of aerobic and anoxic conditions which resulted from the restricted oxygen input and mixing pattern. It was observed that excellent phosphorus removal could be achieved if first all of the aerators were shut off for 12 hr per day for two to three weeks. After this, the period of time when all aerators were off was gradually reduced until excellent phosphorus removal was achieved. The investigators observed very little phosphorus release under anaerobic

conditions in the bottom of the clarifiers, or when glucose, a form of readily biodegradable soluble substrate, was added to samples of the mixed liquor. These observations led to the conclusion that phosphorus was removed by extracellular chemical precipitation rather than by intracellular uptake.

This conclusion is supported somewhat by the work of Arvin and Holm Kristensen (1984) who studied enhanced phosphorus removal using a soft wastewater with an alkalinity of approximately 100 mg/L as $CaCO_3$. They observed that under anaerobic conditions, potassium and magnesium were released along with the phosphorus while calcium was taken up by the sludge. In the two plants that were investigated, 50 to 60 percent of the phosphorus in the sludge was in the form of a chemical precipitate, with calcium and magnesium being the predominant counterions. Arvin and Holm Kristensen (1983) have also postulated that under conditions of low alkalinity, chemical precipitation of phosphorus can occur in fixed biofilms and possibly in bacterial flocs as a result of changes in pH within the film or floc due to denitrification.

A/O PROCESS

The A/O process for biological phosphorus removal is shown in Fig. 20. It was developed in the U.S. by Air Products and Chemicals, Inc. (Hong et al., 1982), and is similar to the Phoredox concept put forth by Barnard (1976) except that the anaerobic and aerobic stages are divided into a number of compartments. It is said that this configuration can be designed with or without nitrification (Hong et al., 1982).
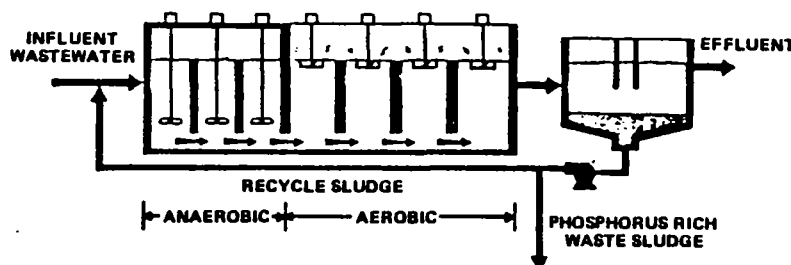

FIG. 20. A/O PROCESS
(From Hong et al., 1982)

If total nitrogen removal is required, a three stage A/O process incorporating an anoxic zone for denitrification is available as shown in Fig. 21. This is similar to the three stage process previously identified as Modified Phoredox except that the anaerobic, anoxic and aerobic zones are divided into a number of compartments.
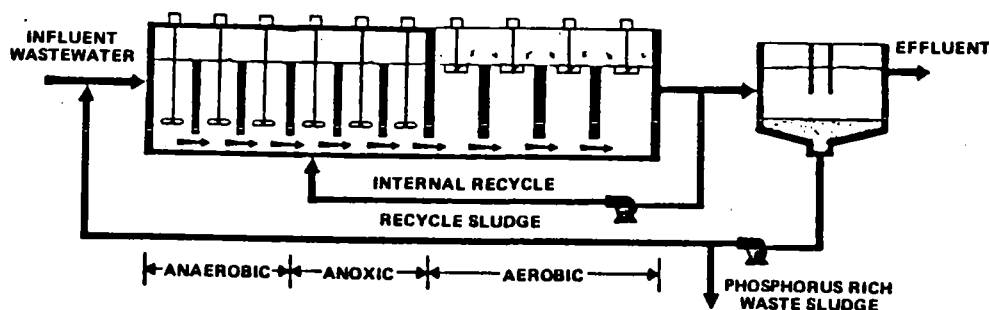
FIG. 21.   THREE STAGE A/O PROCESS
(From Hong et al., 1982)

Typical nominal detention times for the two and three stage A/O processes are given in Table 6 (Hong et al., 1982).

TABLE 6.   NOMINAL DETENTION TIMES FOR A/O PROCESSES

Zone	Detention Time (hr)
Anaerobic	0.5 - 1.0
Anoxic (three stage only)	0.5 - 1.0
Aerobic (nitrification)	3.5 - 6.0
Aerobic (no nitrification)	1.8 - 2.5

The A/O process has been demonstrated at pilot-scale at Rochester, New York; Largo, Florida; Bath, Pennsylvania and Baltimore, Maryland (Hong et al., 1982; Deakyne et al., 1984). In addition, two full-scale treatment plants in the U.S. have been converted to the A/O process. These are at Pontiac, Michigan and Largo, Florida. The experiences at Pontiac are described in another paper at this seminar.

Largo, Florida A/O Plant
The treatment plant at Largo, Florida originally consisted of three parallel contact stabilization activated sludge units with a total capacity of 34 000 m³/d. One of the units was retrofitted to the three stage A/O process utilizing three compartments in the anaerobic zone, two compartments in the anoxic zone, and five compartments in the aerobic zone. The nominal detention times were 0.9 hr in the anaerobic zone, 0.6 hr in the anoxic zone and 2.6 hr in the aerobic zone. Primary clarification was employed. Waste activated sludge was digested aerobically and gravity thickened to about 4 percent solids prior to dewatering on belt filter presses. It was then dried and pelletized and sold as fertilizer.

During the period from September 1980 to February 1982, the effluent total phosphorus concentration averaged 1.8 mg/L with maximum and minimum monthly averages of 3.1 and 1.2 mg/L respectively (Hong et al., 1982). A total phosphorus concentration in the raw wastewater of 8.9 mg/L was given as being typical, and thus the average phosphorus removal was 80 percent. Based on these results, it was decided to also convert the other two contact stabilization units to the A/O process, and add effluent filtration.

PHOSTRIP PROCESS

The PhoStrip process was developed in the U.S. by Levin and co-workers in the late 1960's and early 1970's (Levin and Shapiro, 1965; Shapiro et al., 1967; Levin et al., 1972) and is now marketed by Biospherics Incorporated. The basic flow scheme for this process is shown in Fig. 22. Some portion of the return sludge is exposed to anaerobic conditions in a reactor which is known as the stripper tank and located in a line parallel to the return sludge line. In this tank, phosphorus is released from the sludge into the supernatant. The phosphorus depleted sludge is then returned to the aeration basin where it takes up the phosphorus entering the plant in the influent wastewater. Thus in the PhoStrip process the sludge micro-organisms are exposed to alternating anaerobic and aerobic environments as in the Modified Bardenpho and A/O processes. The main difference is that in the latter processes the return sludge is exposed to anaerobic conditions in the presence of the influent wastewater while in the PhoStrip process the return sludge is exposed to anaerobic conditions on its own. This can be seen as a disadvantage as the presence of soluble substrate stimulates phosphorus release under anaerobic conditions. As a result, it has been proposed that primary effluent be used for elutriation of the stripper tank (Levin et al., 1982). On the other hand, in the PhoStrip process the basin in which an anaerobic environment is to be maintained is protected from hydraulic shock loads and variations in influent quality as it is located in a sidestream and not at the head end of the plant.
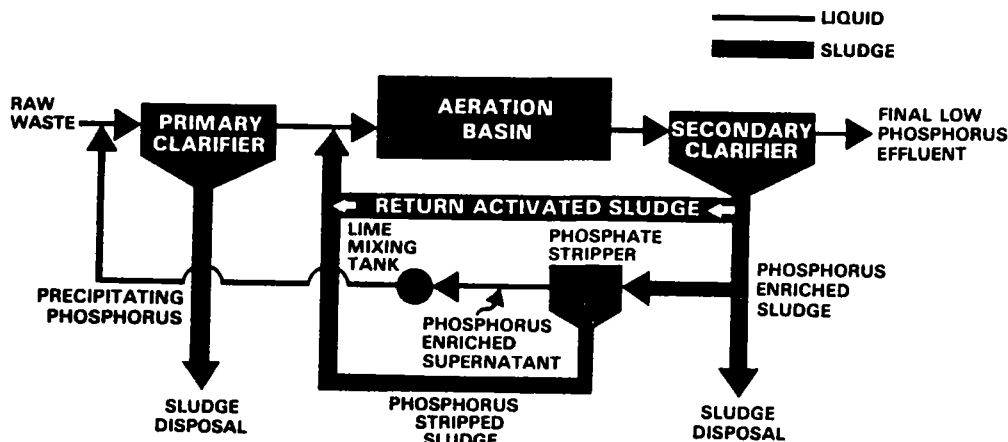

FIG. 22. PHOSTRIP PROCESS
(From Levin et al., 1982)

The phosphorus rich supernatant from the stripper tank is treated with lime and discharged to the primary clarifier as shown in Fig. 22 or can undergo solids separation in a separate small clarifier after lime treatment as shown in Fig. 23. In some cases it may be possible to apply the supernatant directly to land without first undergoing chemical precipitation. If 50 percent of the return sludge flow is passed through the stripper tank, a hydraulic detention time of 8 to 12 hr is normally used for this reactor. These figures are based on the return sludge flowrate being about 30 percent of the influent flowrate. Along with removal by chemical precipitation of the stripper tank supernatant, phosphorus is also removed in the waste activated sludge as in the Modified Bardenpho and A/O processes.
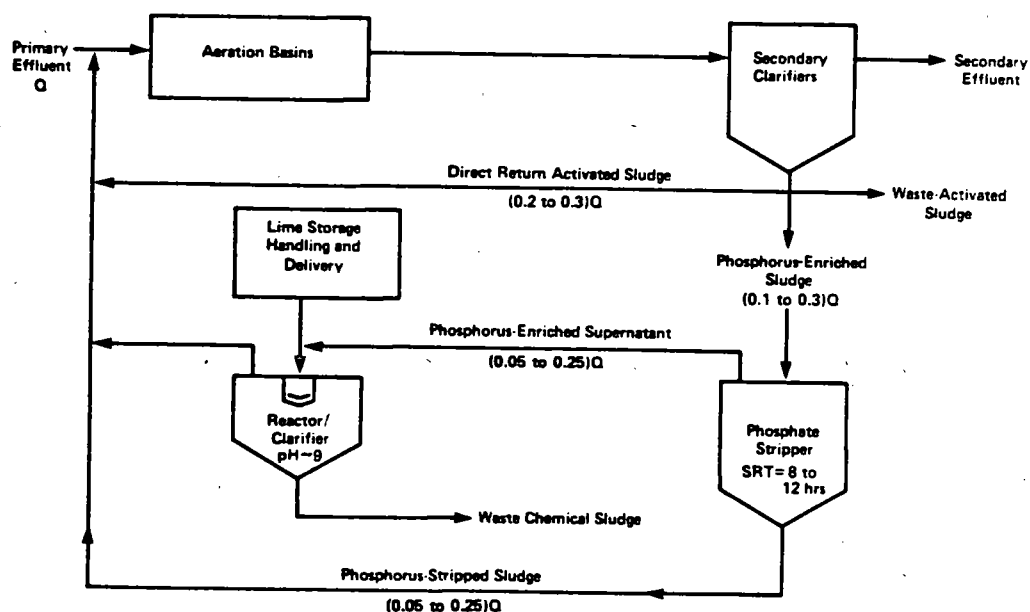
FIG. 23. PHOSTRIP PROCESS
(From Levin et al., 1982)

The PhoStrip process has been implemented at full-scale at fourteen locations in the U.S. The basic design information for these plants is given in Table 7. Performance data for eight of these is summarized in Table 8. Effluent total phosphorus concentrations averaged below 1.0 mg/L as P for five of the plants. It should be noted, however, that this was in each case for a period of operation of less than one year.

TABLE 7. BASIC DESIGN INFORMATION FOR FULL-SCALE PHOSTRIP PLANTS

(From Biospherics Incorporated, 1983)

	Seneca Falls, NY	Lansdale, PA	Lititz, PA	Carpenters-ville, IL	Adrian, MI	Tahoe/ Truckee, NV[a]	Texas City, TX	Ithaca, NY[a]	Savage, MD	Brockton, MA	South-towns, NY	Rochester, MN[a]	Amherst, NY	Reno, NV
Design Flow (MGD)	0.9	2.5	3.5	5.0	7.0	7.4	7.5	10.0	15.0	18.0	16.0	19.1	24.0	30.0
Final Effluent TP Standard (mg/l)	1.0	2.0	1.0	1.0	1.0	1.0	1.0	1.0	0.3[b]	1.0	1.0	1.0	1.0	0.5[b]
Aeration by Oxygen or Air	A	A	A	A	A	0	0	A	A	A	0	0	0	A
Aeration Mode[c]	Comp.	Plug	Plug	Cont.	Conv.	Plug	Plug	Conv.	Plug or Step Feed	Var.	Plug	Comp.	High	Plug
1 or 2 Stage Secondary Treatment	1	2	2	1	2	1	1	1	2	2	1	2	2	1
Equalization	No	Yes	No	No	Yes	Yes	No	No	Yes	No	No	Yes	Yes	No
Final Filtration	No	No	No	No	Yes	Yes	Yes	No	Yes	Yes	Yes	No	Yes	Planned
Sludge Handling[d]	Thick. Anaer.	Thick. Filt.	Thick. Aerob.	Aerob.	Thick. Anaer.	Thick. Anaer.	Press Aerob.	Thick. Anaer.	Thick. Aerob.	Thick. Anaer.	Press Incin.	Aerob. Anaer.	DAF Thick.	Anaer.
Strippers (No.)	1	1	1	1	1	2	2	2	2	1	4	2	2	5
Reactor/Clarifiers (No.) or Mixer/Flocculators (No.)	MF1	RC1	RC1	MF1	MF1	MF2	RC2	MF2	RC2	RC1	RC4	RC2	RC1	MF2
Elutriation Source[e]	SR	RC	RC	SR	Prl.	Prl.	RC Prl.	SR Prl.	RC	RC	RC	SR RC Prl.	RC	SR

a Under construction or in design

b With final filtration; chemical polishing available but not utilized

c Complete mix; Conventional; Contact; Variable; High rate

d Thickening; Aerobic digestion; Anaerobic digestion; Filtration; Filter Press; Incineration; Dissolved Air Flotation

e Sludge Recycle elutriation; Reactor/Clarifier overflow elutriation; Primary effluent supplement

TABLE 8.   PERFORMANCE DATA FOR FULL-SCALE PHOSTRIP PLANTS
(From Levin, 1983)

City	Consulting Engineer	Flow (MGD)	Data Period (Mo.)	Total Phosphorus (mg/l as P)						Notes
				Influent Averages			Effluent Averages			
				Min. Mo.	Avg. Mo.	Max. Mo.	Min. Mo.	Avg. Mo.	Max. Mo.	
Adrian, Michigan	McNamee, Porter & Seeley, Ann Arbor, Michigan	7	11	3.4	4.4	5.5	<0.1	0.4	0.6	a
Amherst, New York	Charles R. Velzy Assoc., Buffalo, New York	22	12	2.9	5.2	14.3	0.4	1.3	2.5	b
Carpentersville, Illinois	Warren & Van Praag, Chicago, Illinois	5	5	3.3	4.5	5.8	0.5	0.5	0.5	c, d
Lansdale, Pennsylvania	Betz, Converse, Murdoch, Plymouth Meeting, Pennsylvania	2.5	12	4.0	5.2	6.4	0.6	1.2	2.0	b
Reno/Sparks, Nevada	Kennedy/Jenks San Francisco, California	6	10	6.0	7.0	8.0	0.3	0.9	2.0	a
		12	1	9.1	9.1	9.1	0.9	0.9	0.9	
		30	4	6.9	7.0	7.1	0.8	1.0	1.0	
Savage, Maryland	Whitman, Requardt & Assoc., Baltimore, Maryland	15	6	5.7	8.1	9.3	0.5	1.2	1.7	b
Seneca Falls, New York	Stearns & Wheler Cazenova, New York	1	1	6.3	6.3	6.3	0.6	0.6	0.6	
Southtowns, New York	URS Engineers Montvale, New Jersey	16	4	2.3	3.2	4.1	0.3	0.5	0.9	

Notes: a. Excludes short periods of unrelated plant upset.
b. Includes process optimization experimental period.
c. "Influent" was primary effluent.
d. P data are for orthophosphate, no total P data taken.

Information on the operation of the plants in Table 8 has been
presented in a proprietary report prepared by Biospherics Incorporated
(1983).   The information in this report is drawn from reports by two U.S.
consulting engineering firms. Additional information on some of these
plants has been given by Walsh et al. (1983). Comments on the operation of
each of these facilities are summarized in the following sections.

Adrian, Michigan
Operation of the PhoStrip process began in June 1981. The effluent
total phosphorus concentration averaged 0.4 mg/L as P over an eleven month
period. It was higher on occasions when upsets in other parts of the plant
had a negative impact on the PhoStrip system. For example, problems with
the sludge digesters required the temporary storage of sludge in the
activated sludge units. As a result, the volume and mass of return sludge
entering the stripper tank were much greater than its design capacity.
Successful phosphorus removal was apparently achieved with a influent BOD_5
concentration as low as 80 mg/L. It is not known if any digester

supernatants returned to the process significantly altered the wastewater characteristics. The PhoStrip system was out of operation for a period of time when there was a mechanical failure of the return sludge pumps.

Amherst, New York

The PhoStrip system began operation in the fall of 1981. The total phosphorus concentration in the effluent averaged 1.3 mg/L as P over a period of one year. There have been a number of mechanical problems with the PhoStrip system. For example, pumps in the lime addition system malfunctioned and there was freezing in the lime slurry lines. In addition, the MLSS concentration in the aeration basins and return sludge flowrate have been significantly higher than anticipated in the design. This was because the activated sludge system was operated as an extended aeration plant in order to reduce sludge disposal costs, and as a result, the stripper tank was overloaded. The PhoStrip system was shut down in March 1983 because two-point ferric chloride addition was found to be more cost-effective for treatment of the low influent phosphorus levels of 3 to 4 mg/L.

Carpentersville, Illinois

Operation of this PhoStrip installation was discontinued after one year because the effluent standard for phosphorus was lifted. The effluent soluble orthophosphate concentration averaged 0.5 mg/L as P over a five month period.

Lansdale, Pennsylvania

The PhoStrip process began operation in April 1981 about five months after startup of a new activated sludge plant. The effluent total phosphorus concentration was on average 1.2 mg/L as P over a twelve month period. Unspecified upsets in the rest of the plant created problems with the operation of the PhoStrip system. Lime scaling in the PhoStrip system also caused significant difficulties. In order to reduce operating costs, the PhoStrip system was shut down in July 1983 and experiments with alum addition were begun.

Reno/Sparks, Nevada

Initially, one and then two of the activated sludge modules at Reno/ Sparks were retrofitted for the PhoStrip process for purposes of process demonstration and evaluation. The data in the first two lines for Reno/ Sparks in Table 8 is for those periods. Few problems were reported to have been encountered . Retrofit of the entire plant for PhoStrip was completed at the end of 1981. Published data for operation of the full plant covers only the first few months of operation and indicates that a total phosphorus effluent standard of 1.0 mg/L as P can be met. The major operational problem was foaming in the aeration basins which had increased significantly since the PhoStrip system was installed. There were also maintenance problems with the lime system.

Savage, Maryland

This PhoStrip system was started up in the summer of 1981 as part of the startup of an entire plant. The effluent total phosphorus concentration averaged 1.2 mg/L as P over a six month period. Startup was

difficult because of equipment problems at the plant. For example, problems with the sludge dewatering equipment necessitated carrying an excessively large sludge inventory in the activated sludge units, and this condition apparently caused an overload of the stripper tank. In addition, anaerobic conditions were not always maintained in the stripper tank because it did not have the capacity to denitrify the nitrate produced in the activated sludge system.

Seneca Falls, New York

A full-scale demonstration was carried out in 1973 at an activated sludge plant retrofitted for the PhoStrip process. During a one month period when the plant was monitored intensively, the effluent total phosphorus concentration was 0.6 mg/L as P on average.

Southtowns, New York

This PhoStrip facility was reported to not have been in continuous service as low influent phosphorus concentrations allowed the effluent standard to be met by normal secondary treatment. An effluent total phosphorus concentration of 0.5 mg/L as P was reported for a four month period.

BIODENIPHO PROCESS

A process for biological nitrogen and phosphorus removal known as the Biodenipho process has been developed in Denmark (Arvin and Holm Kristensen, 1984). It is based on a biological nitrogen removal process known as the Biodenitro process which was also developed in Denmark (Bundgaard et al., 1983).

A basic flow diagram for the Biodenipho process is shown in Fig. 24. The Biodenitro process is the same except that it does not include the initial anaerobic basin. These processes can be described as continuous batch processes with an alternating mode of operation. Nitrification takes place under aerated conditions in a simple basin followed by a period of denitrification under unaerated conditions in the same basin. The advantages of this approach are that it eliminates the need for mixed liquor recycle pumps and a recycle pipe or channel, and the detention times for nitrification and denitrification can be varied as they are not fixed by the volumes of the basins.

Operation of the Biodenipho process is based on four different phases with a typical duration of 1.0 hr each as shown in Fig. 24. Nitrification occurs in basin 2 for 2.0 hrs during Phases A and B and then denitrification of the nitrate produced takes place in this basin for 2.0 hr during Phases C and D. Conversely, nitrification takes place in basin 1 for 2.0 hr during Phases C and D followed by denitrification for 2.0 hr in Phases A and B of the next cycle. During Phases A and C, wastewater does not pass through basins 2 and 1 respectively.

The largest application of the Biodenipho process has apparently been at a 1500 L/d pilot plant at the Technical University of Denmark (Arvin and Holm Kristensen, 1984). The anaerobic basin had a volume of 180 L and was divided into three compartments. The aerobic/anoxic basins had a volume of

Phase A   Phase B

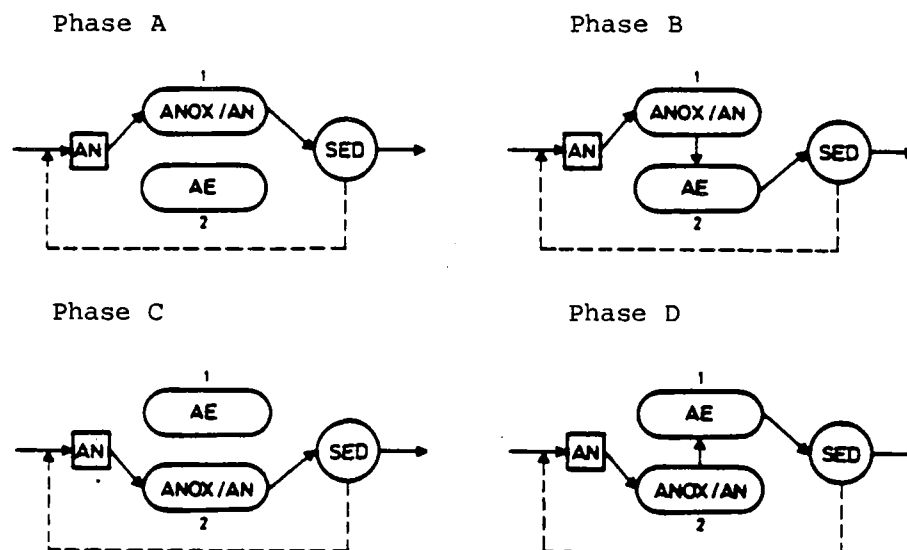Phase C   Phase D

FIG. 24.   BIODENIPHO PROCESS
(From Arvin and Holm Kristensen, 1984)

800 L each. The total phosphorus concentration was 10 to 12 mg/L as P in
the influent and 4 to 5 mg/L as P in the effluent. Approximately 60
percent of the phosphorus in the sludge mass was found to be in the form of
a chemical precipitate.

There are about ten full-scale Biodenitro plants in Denmark. Enhanced
phosphorus removal without the addition of chemicals has been observed at
several of these, presumably after anaerobic conditions were created in
anoxic basins (Bundgaard et al., 1983). At Frederikssund and Faaborg,
total phosphorus concentrations of about 11 mg/L as P in the influent were
reduced to about 3 mg/L as P in the effluent. At the Frederikssund plant,
approximately 50 percent of the phosphorus in the sludge was found to be in
the form of a chemical precipitate.

COMPARISON OF A/O, BARDENPHO AND PHOSTRIP PROCESSES

Walsh et al. (1983) have compared the two stage A/O, Modified
Bardenpho and PhoStrip processes by ranking them subjectively against six
criteria as shown in Table 9. The main advantage of the two stage A/O
process was seen to be its simplicity. Superior dependability, flexibility
and ease of control as a result of its relatively long hydraulic detention
time were considered to be the strengths of the Modified Bardenpho
process. The advantage of the PhoStrip process was thought to be that the
anaerobic reactor is protected from hydraulic shock loads and variations in
influent quality because of its location in a sidestream rather than at the
head of the plant. The main disadvantage of the PhoStrip process was
considered to be its operational complexity as a result of the stripper
tank, lime addition, lime clarifier and need for elutriation, and the
impact of variations in operating conditions in the rest of the plant on
the PhoStrip system.

TABLE 9. SUBJECTIVE RANKING OF BIOLOGICAL PHOSPHORUS REMOVAL PROCESSES*

Criteria	Two stage A/O	Modified Bardenpho	PhoStrip
1. Response to hydraulic surges	3	2	1
2. Response to variation in waste strength	3	2	1
3. General dependability	3	1	2
4. Controllability	3	1	2
5. Flexibility	3	1	2
6. Complexity	1	2	3

* 1 is most favourable, 3 is least favourable.

In addition to its simplicity, another advantage of the two stage A/O process is that the anaerobic stage is divided into several compartments. As a result, for a given concentration of readily biodegradable soluble substrate in the influent wastewater, the concentration of readily biodegradable soluble substrate in the first compartment is higher than that if the substrate was discharged to an anaerobic stage of the same volume but without compartments. This is because of the smaller volume for dilution in the first compartment. Initial exposure of the micro-organisms to the higher concentration of readily biodegradable soluble substrate in the first compartment might better stimulate phosphorus release at times when the soluble substrate concentration in the influent wastewater is low. However, a significant disadvantage of the two stage A/O process is the potential for nitrate formed in the aerobic stage to be directly discharged to the anaerobic stage via the return sludge. This is of serious concern because practical experience has shown that under typical operating conditions, it is virtually impossible to completely eliminate nitrification during at least the summer months, and possibly year round (Hill, 1985).

CLOSING REMARKS

A number of different processes for enhanced biological phosphorus removal have been described. In each of these processes, phosphorus release under anaerobic conditions and phosphorus uptake under aerobic conditions are the two basic steps. In order for biological phosphorus removal to occur, dissolved oxygen and nitrate must be absent from the anaerobic reactor and readily biodegradable soluble substrate must be present.

The success of these processes in consistently meeting a total phosphorus effluent standard of 1.0 mg/L as P under pilot and full-scale operation has been variable. In some cases it is difficult to determine the exact causes of poor performance because the available published reports do not contain adequate information. In many ways, these processes are still in a developmental stage. It is difficult to make a final judgement on aspects such as reliability because many of the full-scale applications have been in operation for only a short period of time, and innovations in the design and operation of the processes are still being developed.

The costs of the biological phosphorus removal processes relative to each other and to chemical phosphorus removal have not been addressed in this paper. This important aspect is discussed in another paper at this seminar.

REFERENCES

Arvin, E., and G. Holm Kristensen (1983). Phosphate precipitation in biofilms and flocs. Wat. Sci. Tech., 15, 3/4, 65-85.

Arvin, E., and G. Holm Kristensen (1984). Exchange of organics, phosphate, and cations between sludge and water in biological phosphorus and nitrogen removal processes. Presented at IAWPRC Post-conference Seminar on Enhanced Biological Phosphorus Removal from Wastewater, Paris, France.

Bargman, R.D., J.M. Betz and W.F. Garber (1971). Nitrogen-phosphate relationships and removals obtained by treatment process at the Hyperion Treatment Plant. Advances in Wat. Poll. Res., 1, I-14/1 - I-14/17.

Barnard, J.L. (1974). Cut P and N without chemicals. Wat. & Wastes Engng, 11, 7, 33-36.

Barnard, J.L. (1976). A review of biological phosphorus removal in the activated sludge process. Water S.A., 2, 136-144.

Barth, E.F., R.C. Brenner and R.F. Lewis (1968). Chemical-biological control of nitrogen and phosphorus in wastewater effluent. Jour. Wat. Poll. Cont. Fed., 40, 2040-2054.

Biospherics Incorporated (1983). PhoStrip status report. Biospherics Inc., 4928 Wyaconda Rd., Rockville, MD 20852.

Bundgaard, E., G. Holm Kristensen and E. Arvin (1983). Full-scale experience with phosphorus removal in an alternating system. Wat. Sci. Tech., 15, 3/4, 197-217.

Burdick, C.R., D.R. Refling and H.D. Stensel (1982). Advanced biological treatment to achieve nutrient removal. Jour. Wat. Poll. Cont. Fed., 54, 1078-1086.

Comeau, Y. (1984). Biochemical Models for Biological Excess Phosphorus Removal from Wastewater. M.A.Sc. Thesis, Department of Civil Engineering, University of British Columbia.

Deakyne, C.W., M.A. Patel and D.J. Krichten (1984). Pilot plant demonstration of biological phosphorus removal. Jour. Wat. Poll. Cont. Fed., 56, 867-873.

Di Gregorio, D. (1985). Personal communication.

Ekama, G.A., and G.v.R. Marais (1984a). "Nitrification" in Theory, Design and Operation of Nutrient Removal Activated Sludge Processes. Water Research Commission, Pretoria, South Africa.

Ekama, G.A., and G.v.R. Marais (1984b). "Biological Nitrogen Removal" in Theory, Design and Operation of Nutrient Removal Activated Sludge Processes. Water Research Commission, Pretoria, South Africa.

Ekama, G.A., G.v.R. Marais and I.P. Siebritz (1984). "Biological Excess Phosphorus Removal" in Theory, Design and Operation of Nutrient Removal Activated Sludge Processes. Water Research Commission, Pretoria, South Africa.

Evans, B.W. (1983). Biochemical Transformations Involved in Phosphorus Release and Uptake from Wastewater by Activated Sludge. M.A.Sc. Thesis, Department of Civil Engineering, University of Toronto.

Florentz, M., and P. Hartemann (1984). Screening for phosphate accumulating bacteria isolated from activated sludge. Environmental Technology Letters, 5, 457-463.

Gerber, A., M.J. Simpkins, C.T. Winter and J.A. Scheepers (1982). Biological Nutrient Removal from Wastewater Effluents: Performance Evaluation of the Phoredox and UCT Processes. Contract Report C WAT 52, Council for Scientific and Industrial Research, Pretoria, South Africa.

Grady, C.P.L., and H.C. Lim (1980). Biological Wastewater Treatment: Theory and Applications. Marcel Dekker, Inc., New York.

Hill, R.D. (1985). Personal communication.

Hong, S.-N., D.J. Krichten, K.S. Kisenbauer and R.L. Sell (1982). A biological wastewater treatment system for nutrient removal. Presented at EPA Workshop on Biological Phosphorus Removal in Municipal Wastewater Treatment, Annapolis, Maryland.

Irvine, R.L., H.D. Stensel and J.E. Alleman (1982). Summary Report: Workshop on Biological Phosphorus Removal in Municipal Wastewater Treatment. U.S. Environmental Protection Agency, Cincinnati, Ohio.

Kerdachi, D.A., and M.R. Roberts (1983). Full-scale phosphate removal experiences in the Umhlatuzana Works at different sludge ages. Wat. Sci. Tech., 15, 3/4, 261-281.

Levin, G.V. (1983). PhoStrip process. Presented at EPA Workshop on Biological Phosphorus Removal in Municipal Wastewater Treatment, San Francisco, California.

Levin, G.V., D.M. Maase and T. Kish (1982). The PhoStrip process for biological removal of phosphorus from wastewater. Presented at EPA Workshop on Biological Phosphorus Removal in Municipal Wastewater Treatment, Annapolis, Maryland.

Levin, G.V., and J. Shapiro (1965). Metabolic uptake of phosphorus by wastewater organisms. Jour. Wat. Poll. Cont. Fed., 37, 800-821.

Levin, G.V., G.J. Topol and A.G. Tarnay (1975). Operation of full-scale biological phosphorus removal plant. Jour. Wat. Poll. Cont. Fed., 47, 577-590.

Levin, G.V., G.J. Topol, A.G. Tarnay and R.B. Samworth (1972). Pilot plant tests of a phosphate removal process. Jour. Wat. Poll. Cont. Fed., 44, 1940-1954.

Ludzack, F.J. and M.B. Ettinger (1962). Controlling operation to minimize activated sludge effluent nitrogen. Jour. Wat. Poll. Cont. Fed., 34, 920-931.

Milbury, W.F., D. McCauley and C.H. Hawthorne (1971). Operation of conventional activated sludge for maximum phosphorus removal. Jour. Wat. Poll. Cont. Fed., 43, 1890-1901.

Nicholls, H.A. (1975). Full-scale experimentation on the new Johannesburg extended aeration plants. Water SA, 1, 121-132.

Nicholls, H.A., A.R. Pitman and D.W. Osborn (1984). The readily biodegradable fraction of sewage: its influence on phosphorus removal and measurement. Presented at IAWPRC Post-conference Seminar on Enhanced Biological Phosphorus Removal from Wastewater, Paris, France.

Osborn, D.W., and H.A. Nicholls (1978). Optimization of the activated sludge process for the biological removal of phosphorus. Prog. Wat. Tech., 10, 261-277.

Osborn, D.W., A.R. Pitman and S.L. Venter (1979). Design and operating experience with nutrient removing activated sludge plants in Johannesburg. In Nutrient Removal from Municipal Effluents, Water Research Commission, Pretoria, South Africa.

Paepcke, B.H. (1983). Performance and operational aspects of biological phosphate removal plants in South Africa. Wat. Sci. Tech., 15, 3/4, 219-232.

Shapiro, J., G.V. Levin and Z.G. Humberto (1967). Anoxically induced release of phosphate in wastewater treatment. Jour. Wat. Poll. Cont. Fed., 39, 1810-1818.

Siebritz, I.P., G.A. Ekama and G.v.R. Marais (1983). A parametric model for biological excess phosphorus removal. Wat. Sci. Tech., 15, 3/4, 127-152.

Vacker, D., C.H. Connell and W.N. Wells (1967). Phosphate removal through municipal wastewater treatment at San Antonio, Texas. Jour. Wat. Poll. Cont. Fed., 39, 750-771.

Venter, S.L., J. Halliday and A.R. Pitman (1978). Optimization of the Johannesburg Olifantsvlei extended aeration plant for phosphorus removal. Prog. Wat. Tech., 10, 279-292.

Walsh, T.K., B.W. Behrman, E.R. Jones, and G.W. Weil (1983). A review of biological phosphorus removal technology. Presented at Annual Conference of the Water Pollution Control Federation, Atlanta, Georgia.

Water Pollution Control Federation (1977). Wastewater Treatment Plant Design. Water Pollution Control Federation, Washington, D.C.

Wuhrmann, K. (1964). Nitrogen removal in sewage treatment process. Vehr. Int. Ver. Theor. Angew. Limnol., 15, 580 (in German) (as seen in Ekama and Marais, 1984b).

TECHNICAL AND ECONOMICAL FEASIBILITY OF RETROFITTING

EXISTING MUNICIPAL WASTEWATER TREATMENT PLANTS

IN CANADA FOR BIOLOGICAL PHOSPHORUS REMOVAL

Stephen G. Nutt

Canviro Consultants Ltd.

Kitchener, Ontario

and

N. W. Schmidtke

Norbert W. Schmidtke & Assoc. Ltd.

Kitchener, Ontario

TECHNICAL AND ECONOMICAL FEASIBILITY OF RETROFITTING EXISTING MUNICIPAL WASTEWATER TREATMENT PLANTS IN CANADA FOR BIOLOGICAL PHOSPHORUS REMOVAL

Stephen G. Nutt
Canviro Consultants Ltd.
Kitchener, Ontario

and

N. W. Schmidtke
Norbert W. Schmidtke & Assoc. Ltd.
Kitchener, Ontario

INTRODUCTION AND OBJECTIVES

Phosphorus removal has been practised at most municipal wastewater treatment plants in Canada and the United States discharging to the Lower Great Lakes Basin since the late 1970's. In other areas where receiving waters have been shown to be sensitive to nutrient load, phosphorus removal requirements have been implemented on a site-specific basis. In the case of nitrogenous compounds, the requirement to achieve control of ammonia and/or oxidized nitrogen compounds (nitrate and nitrite) has been site-specific.

Traditionally, phosphorus removal strategies have incorporated chemical precipitation by the addition of metal salts (alum, ferric chloride, lime). This technology is suitable for achieving effluent discharge levels of approximately 1 mg/L total phosphorus; however, cost factors increase significantly when more stringent requirements are placed on the effluent total phosphorus concentrations. Furthermore, chemical precipitation alone does not provide for control of nitrogenous compounds in the discharges.

Several new wastewater process treatment trains have been developed to biologically achieve a high degree of phosphorus removal. In some instances, complete control of nutrient discharges (nitrogen and phosphorus) can be achieved using these configurations. These processes, although relatively unproven in Canada, appear to offer advantages relative to the conventional chemical precipitation technology. However, the technologies are complex compared to the conventional practice. As well, the economic benefits, particularly in retrofit situations, can be very site-specific.

In order to evaluate the technical and economic feasibility of retrofitting existing municipal wastewater treatment plants in Canada for biological nutrient control, Environment Canada's Wastewater Technology Centre retained Canviro Consultants Ltd. to compile technical and cost information necessary to identify the type and size of existing treatment plants in Canada which are most suited to the application of these biological nutrient removal technologies. The specific objectives of the program were:

i) to examine the technical and economic feasibility of retrofitting model primary and secondary (conventional activated sludge and extended aeration) wastewater treatment plants for biological phosphorus removal relative to the costs associated with chemical phosphorus removal, and

ii) to develop a methodology, including an accessories list with cost factors for the various components required to retrofit existing plants, which could be utilized by design engineers to estimate the retrofit costs associated with a specific wastewater treatment plant installation.

It should be noted that the scope of the economic assessment conducted was not limited to a single specific plant, but was conducted in a broader context in order to identify the conditions under which chemical or biological phosphorus removal was more cost effective.

SCOPE OF WORK

In order to streamline the investigation and reduce the number of retrofit situations requiring cost analysis to a manageable level, the scope of work was limited to the following situations:

i) Three types of plants were to be considered for retrofitting to biological phosphorus removal processes, including primary plants, conventional activated sludge plants and extended aeration plants. High-rate activated sludge plants were eliminated from the scope of work due to the limited number of examples of this technology in Canada.

ii) The biological nutrient control technologies to be evaluated were to include the three commercially available processes presently used in North America (Bardenpho licensed to the Eimco Division of Envirotech Corporation; Phostrip licensed to Biospherics Inc.; and A/O licensed to Air Products and Chemicals Inc.) as well as the developmental UCT process.

iii) Ferric chloride was used for the purposes of developing comparative costs for the chemical precipitation technology due to its widespread use in Ontario. A sensitivity analysis was conducted to determine the impact of chemical cost and chemical type on the cost data.

iv) Two total phosphorus discharge limits were imposed on the effluents from the treatment facilities (1.0 mg/L and 0.3 mg/L TP). However, as it was not considered technically feasible to achieve the lower limit (0.3 mg/L) by chemical addition to a primary plant, this cost analysis was done only to a limit of 1.0 mg/L TP. Table 1 summarizes the retrofit situations included in the economic assessment.

v) The design criteria for each type of plant which was to be subjected to retrofitting for phosphorus removal were based on the Ontario Ministry of the Environment design guidelines[1]. Each plant type was evaluated at three levels of flow. The flowrates selected were the lower quartile, median and upper quartile flows based on frequency distributions of flowrates for each type of plant in Ontario. The plant sizes included in the cost analysis are presented in Table 2.

TABLE 1. SUMMARY OF PLANT TYPES INCLUDED IN COST ANALYSIS

MODEL PLANT TYPE	RETROFIT TECHNOLOGY	EFFLUENT TP LIMIT (mg/L)
Primary	Chemical Bardenpho UCT A/O Phostrip	1.0 1.0, 0.3 1.0, 0.3 1.0, 0.3 1.0, 0.3
Conventional Activated Sludge	Chemical Bardenpho UCT A/O Phostrip	1.0, 0.3 1.0, 0.3 1.0, 0.3 1.0, 0.3 1.0, 0.3
Extended Aeration	Chemical Bardenpho UCT A/O Phostrip	1.0, 0.3 1.0, 0.3 1.0, 0.3 1.0, 0.3 1.0, 0.3

TABLE 2. SIZES OF TREATMENT PLANTS EXAMINED FOR PROCESS RETROFIT

MODEL PLANT TYPE	FLOW (Ml/d)		
	LOWER LEVEL	MIDRANGE LEVEL	UPPER LEVEL
Primary	2.3	9.1	36.4
Conventional Activated Sludge	4.5	13.6	36.4
Extended Aeration	0.91	1.6	3.2

vi) The raw wastewater characteristics were as follows:

$$COD = 250 \text{ mg/L}$$
$$TKN = 25 \text{ mg/L}$$
$$Alkalinity = 240 \text{ mg/L}$$
$$Total\ P = 5 \text{ mg/L}$$

The COD:TKN ratio (10:1) influenced the design of the biological nutrient control processes as discussed in Section 3.

METHODOLOGY

Design and Cost Analysis of Model Plants

Process flowsheets for each type of plant which was to be retrofitted for phosphorus removal are presented in Figure 1 (primary plant), Figure 2 (conventional activated sludge plant) and Figure 3 (extended aeration plant), along with the most pertinent process design criteria. It is important to note that these plant designs represent idealized configurations, although standard design practices were followed wherever possible. The major exception to these standard design practices was in the layout of the extended aeration plants. In the smaller size ranges evaluated, packaged treatment plant configurations are typically used. These package plants are not well suited to retrofitting to biological phosphorus removal processes. Therefore, the smaller extended aeration plants were designed as standard mechanical plants with separate components rather than as package plants for the purposes of the cost assessment. The retrofit costs of the biological nutrient control technologies will be influenced by the actual plant equipment and the plant layout at a specific location.
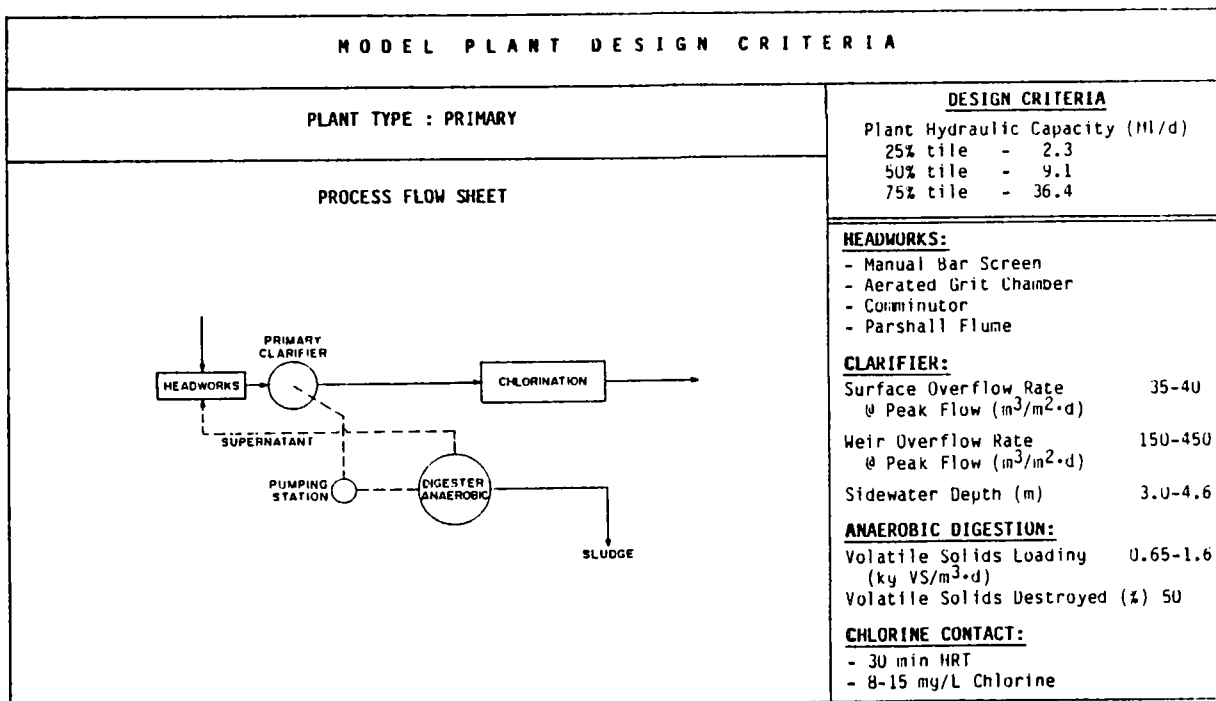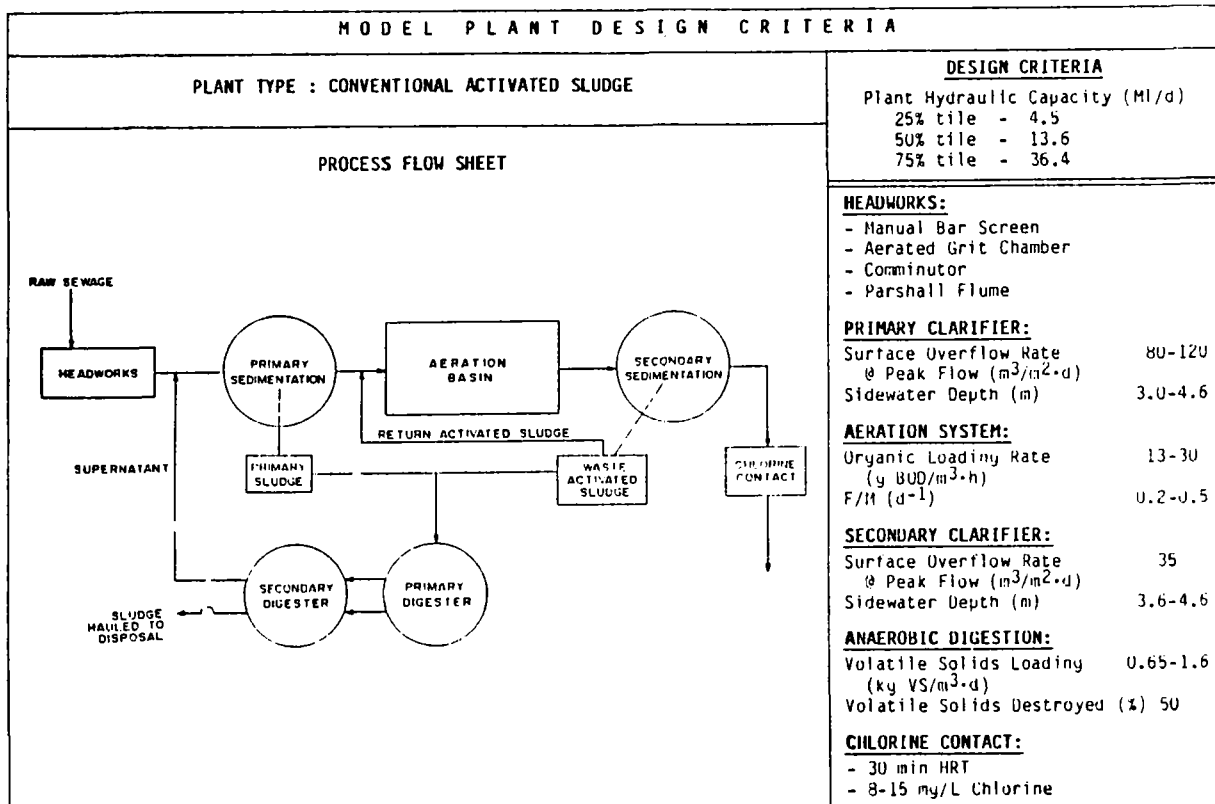

MODEL PLANT DESIGN CRITERIA

PLANT TYPE : PRIMARY

PROCESS FLOW SHEET

DESIGN CRITERIA
Plant Hydraulic Capacity (Ml/d)
25% tile - 2.3
50% tile - 9.1
75% tile - 36.4

HEADWORKS:
- Manual Bar Screen
- Aerated Grit Chamber
- Comminutor
- Parshall Flume

CLARIFIER:
Surface Overflow Rate @ Peak Flow ($m^3/m^2 \cdot d$) — 35-40
Weir Overflow Rate @ Peak Flow ($m^3/m^2 \cdot d$) — 150-450
Sidewater Depth (m) — 3.0-4.6

ANAEROBIC DIGESTION:
Volatile Solids Loading ($kg\ VS/m^3 \cdot d$) — 0.65-1.6
Volatile Solids Destroyed (%) 50

CHLORINE CONTACT:
- 30 min HRT
- 8-15 mg/L Chlorine

FIGURE 1: PRIMARY PLANT PROCESS FLOWSHEET

MODEL PLANT DESIGN CRITERIA

PLANT TYPE : CONVENTIONAL ACTIVATED SLUDGE

DESIGN CRITERIA

Plant Hydraulic Capacity (Ml/d)
- 25% tile - 4.5
- 50% tile - 13.6
- 75% tile - 36.4

PROCESS FLOW SHEET

HEADWORKS:
- Manual Bar Screen
- Aerated Grit Chamber
- Comminutor
- Parshall Flume

PRIMARY CLARIFIER:

Surface Overflow Rate @ Peak Flow (m³/m²·d)	80-120
Sidewater Depth (m)	3.0-4.6

AERATION SYSTEM:

Organic Loading Rate (g BOD/m³·h)	13-30
F/M (d⁻¹)	0.2-0.5

SECONDARY CLARIFIER:

Surface Overflow Rate @ Peak Flow (m³/m²·d)	35
Sidewater Depth (m)	3.6-4.6

ANAEROBIC DIGESTION:

Volatile Solids Loading (kg VS/m³·d)	0.65-1.6
Volatile Solids Destroyed (%)	50

CHLORINE CONTACT:
- 30 min HRT
- 8-15 mg/L Chlorine

FIGURE 2: CONVENTIONAL ACTIVATED SLUDGE PLANT PROCESS FLOWSHEET

MODEL PLANT DESIGN CRITERIA

PLANT TYPE : EXTENDED AERATION ACTIVATED SLUDGE

DESIGN CRITERIA

Plant Hydraulic Capacity (Ml/d)
- 25% tile - 0.91
- 50% tile - 1.6
- 75% tile - 3.2

PROCESS FLOW SHEET

HEADWORKS:
- Manual Bar Screen
- Aerated Grit Chamber
- Comminutor
- Parshall Flume

AERATION SYSTEM:

Organic Loading Rate @ Peak Flow (m³/m²·d)	7-10
F/M (d⁻¹)	0.05-0.15

SECONDARY CLARIFIER:

Surface Overflow Rate @ Peak Flow (m³/m²·d)	35
Sidewater Depth (m)	3.0-4.6

AEROBIC DIGESTION:

Volatile Solids Loading (kg VS/m³·d)	1.6
Volatile Solids Destroyed (%)	40

CHLORINE CONTACT:
- 30 min HRT
- 8-15 mg/L Chlorine

FIGURE 3: EXTENDED AERATION PLANT PROCESS FLOWSHEET

The capital and operating costs for each model plant were developed using CAPDET, the Computer Assisted Procedure for the Design and Evaluation of Wastewater Treatment Systems, which was developed by the United States Army Corp of Engineers on behalf of the U.S. EPA[2]. This computer model estimates total capital costs, and operation and maintenance (O&M) costs based on process design specifications, wastewater and effluent characteristics and unit cost data which are provided as user input to the model. It has been shown that CAPDET is capable of estimating construction costs to within ± 20 percent of actual costs and O&M costs to within ± 30 percent for Canadian situations provided that care is taken to ensure that process design details and internal recycle streams are maintained within realistic ranges[3].

Design of Biological Nutrient Control Processes

The designs of the biological nutrient control processes were based on a thorough review of present design practices for pilot scale, demonstration scale and full scale examples of these technologies. In addition, the suppliers of each of the commercially available technologies (Bardenpho, A/O and Phostrip) were contacted and asked to provide process design data based on the specified wastewater characteristics and desired effluent TP concentrations. Each of the biological nutrient control technologies included in the technical and economic assessment are described briefly in the following subsections. More detailed descriptions of the processes are available elsewhere[4,5].

Bardenpho. The Bardenpho (Phoredox) process can be configured either as a five-stage or a three-stage process depending on the degree of nitrogen control which must be achieved. For the purposes of this evaluation, a five-stage system was specified. In the five-stage system, as shown schematically in Figure 4, raw wastewater and return activated sludge are introduced to a mixed anaerobic basin. The mixture then flows to a mixed anoxic basin into which a recycle stream from the next downstream aerobic basin is introduced. From the anoxic basin, the mixed liquor flows in sequence to an aerated aerobic basin, a mixed anoxic basin, a final aerated aerobic basin and then to the secondary clarifier. The process objectives of the 5-stage system are high degrees of N and P removal. When less complete N removal can be tolerated, a 3-stage system is appropriate. The flowsheet for this process is identical to the 5-stage system with the exception that the last two stages (one anoxic and one aerobic) are deleted.

Bardenpho and Phoredox systems have been applied almost uniquely to nitrifying systems. However, there is no reason why they should not work in non-nitrifying plants as a two-stage system similar to A/O.

UCT. The UCT process can be considered as an extension of the 3-stage Phoredox process (Figure 5. It differs only in that the return activated sludge stream, rather than being fed directly to the initial anaerobic reactors, is routed through the anoxic reactor, and hence, to the anaerobic basin. This modification provides an additional barrier to the entry of nitrate into the anaerobic basin.
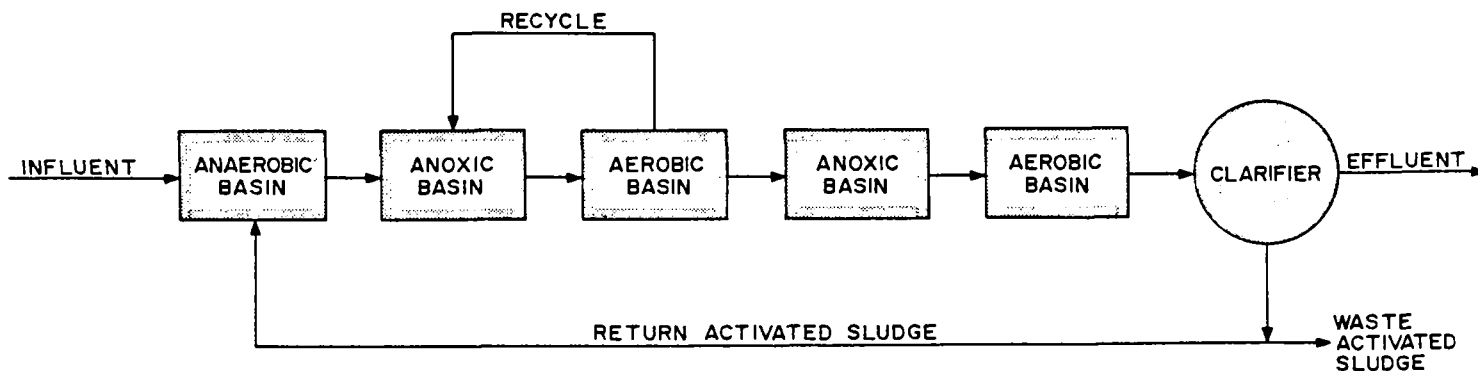
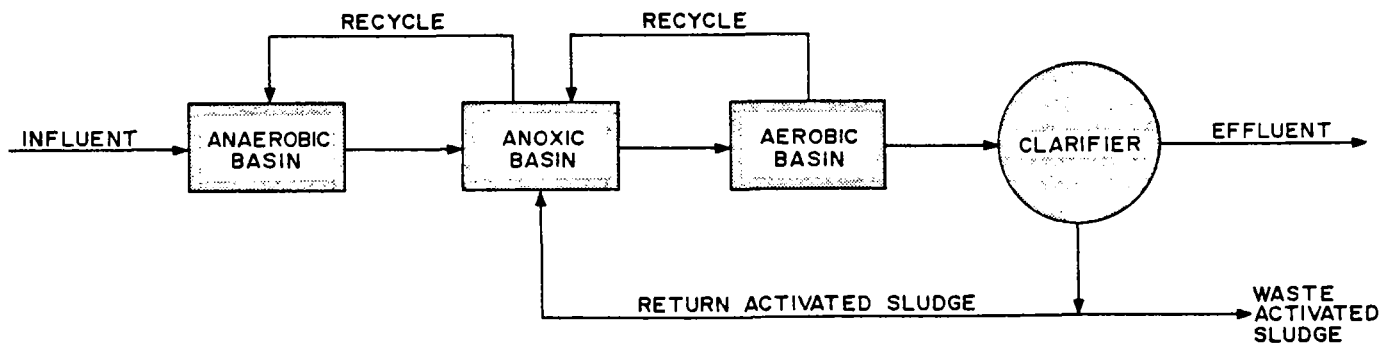FIGURE 4: SCHEMATIC FLOWSHEET OF FIVE-STAGE BARDENPHO PROCESS

FIGURE 5: SCHEMATIC FLOWSHEET OF UCT PROCESS

A/O and A^2/O. The flowsheet of the A^2/O process is identical to that of the 3-stage Bardenpho process. This flowsheet would be applicable to a nitrifying system where partial denitrification is required. In the A/O process (Figure 6), influent waste and returned activated sludge flow first into a mixed anaerobic basin, then to an oxygenated (or aerated) aerobic basin and finally to the secondary clarifier. This flowsheet is applicable to a non-nitrifying system and is identical to the 2-stage Bardenpho flowsheet mentioned previously. In detail, designs of Bardenpho and A/O systems have differed in the degree of compartmentalization of the basins (greater degree in A/O systems) and the use of high purity oxygen (A/O) rather than air (Bardenpho) for aeration. The two-stage A/O system was specified for the purposes of the cost estimation.

Phostrip. The Phostrip process, as shown schematically in Figure 7, is a sidestream process in which the operations critical to phosphorus removal are conducted on the return activated sludge stream rather than on the wastewater stream. The mainstream flowsheet for the process is the same as for a conventional activated sludge plant. The return activated sludge stream is split so that a fraction passes through an anaerobic basin called a phosphorus stripper. Two stripper flowsheets have been used. The recycle stripper removes released phosphorus by recycling stripper supernatant through the sludge blanket. The elutriation stripper, shown in Figure 7, uses primary effluent to strip released phosphorus and is preferred because the primary effluent introduces soluble biodegradable organic matter required for anaerobic phosphorus release from the sludge. Phosphorus is removed from the stripper overflow by precipitation as calcium phosphate using lime. If the activated sludge system is nitrifying, the return activated sludge must first pass through an anoxic, stirred denitrification basin prior to the phosphorus stripper. For this evaluation, an non-nitrifying system was selected with phosphorus release in an elutriation stripper and calcium phosphate precipitation in a separate reactor-clarifier.

Retrofit Design Details. A thorough review of the performance of demonstration scale and full scale biological phosphorus removal systems indicated that the totally biological processes (Bardenpho, A/O and UCT) would require final effluent filtration to consistently achieve an effluent total phosphorus limit of 1 mg/L. Furthermore, in order to consistently achieve the lower specified effluent phosphorus level (0.3 mg/L), these processes would require the supplemental addition of metal salts in conjunction with effluent filtration. Therefore, these processes were incorporated into the process flow schemes. In the case of Phostrip, performance data indicated that an effluent level of 1 mg/L could be achieved without effluent filtration, but filtration would be required to achieve a level of 0.3 mg/L TP.
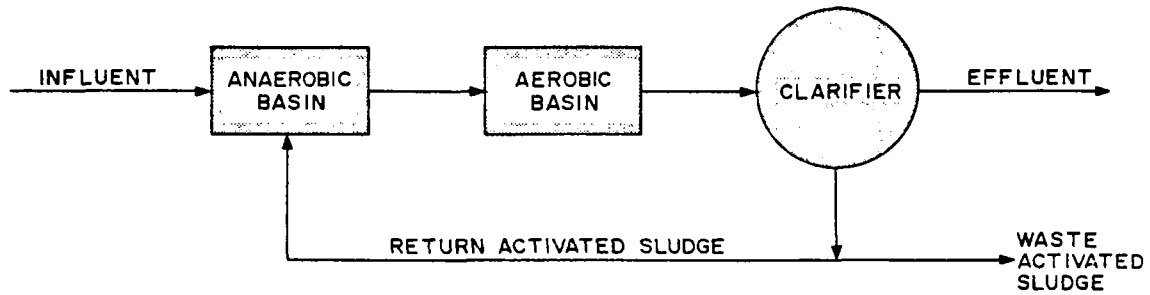
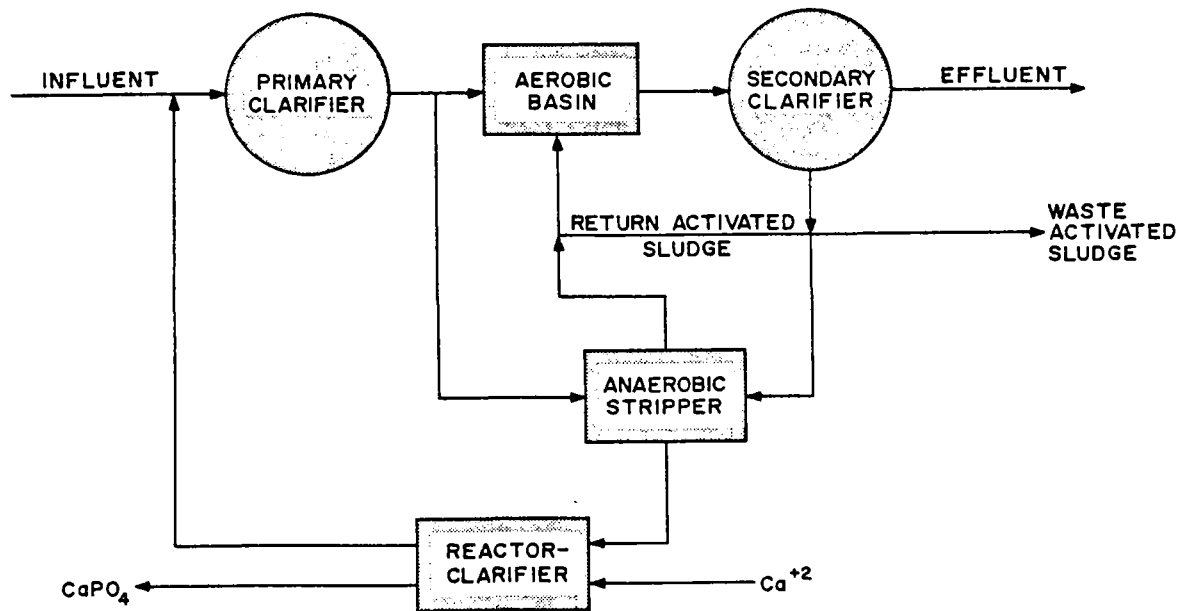FIGURE 6: SCHEMATIC FLOWSHEET OF A/O PROCESS

FIGURE 7: SCHEMATIC FLOWSHEET OF NON-NITRIFYING PHOSTRIP
PROCESS WITH ELUTRIATION STRIPPER

The retrofit designs were intended to incorporate existing equipment and tankage to the greatest extent possible. Therefore, the actual process design details (basin HRTs) varied slightly from case-to-case in order to maximize the use of existing equipment but, at the same time, maintain typical design ranges normally specified for the selected processes. Compartmentalization of the individual basins was practised and flexibility in the use of the basins between anaerobic, anoxic and aerobic environments was designed into the system through the use of submerged turbine aerators.

The characteristics of the raw wastewater were marginal for successful application of the biological phosphorus removal technologies which incorporate nitrogen control (Bardenpho and UCT). It is generally accepted that a minimum COD/TKN ratio of at least 10:1 is required for successful operation[6]. To maximize the COD/TKN ratio, the primary clarifier in the conventional activated sludge plant was bypassed for the retrofit situations.

Sludge handling processes applied will have a significant impact on the performance of the biological phosphorus removal systems. The phosphorus sink is either the biological solids alone or in combination with a calcium phosphate sludge. Anoxic conditions and low pH will cause release of the bound phosphorus from these solids back into solution. Therefore, waste sludge concentration should be by "aerobic" processes such as dissolved air flotation. Anaerobic digester supernatant return to the main-stream biological system should be avoided to prevent recycle of phosphorus. Alternately, the digester supernatant should be treated with metal salts to precipitate the released phosphorus and prevent its return to the main-stream processes. The retrofit designs allowed flexibility in the handling of the sludge streams as the most economic option is very site-specific and is impacted by local guidelines with respect to sludge disposal.

Design of Chemical Precipitation Processes

Simultaneous precipitation of phosphorus by the addition of a metal salt (alum or ferric chloride) to the aeration tank of conventional activated sludge plants or extended aeration plants is widely practised and well-proven in Ontario. Ferric chloride was selected as the metal salt. Chemical dosages of 8 mg Fe^{3+}/L to achieve an effluent total phosphorus limit of 1.0 mg TP/L and 14 mg Fe^{3+}/L to achieve an effluent total phosphorus limit of 0.3 mg TP/L were selected based on experience in Ontario[7] and the United States[8]. A final effluent granular media filter was included in the flow scheme to achieve the lower effluent limit. The increased sludge production rates produced as a result of metal salt addition were based on Ontario experience[9].

Calculation of Costs of Retrofitting for Phosphorus Removal

A list of process equipment or accessories required to retrofit each type of model plant to chemical or biological phosphorus removal was developed. This accessory list included such hardware as pumps, piping, tankage, blowers, mixers, granular media filters and lime feed equipment. Capital and O&M cost curves were developed for each type of equipment for a range of flows up to 40.0 Ml/d. These cost curves were based on equipment vendors' information, Canviro's own cost data and on cost data included in the CAPDET model.

For each retrofit situation, the additional equipment required for the retrofit was selected and costs (capital and O&M) were determined from the pertinent cost curves at the required flow. These retrofit cost items were then added to the model plant costs developed from the CAPDET model. Wherever appropriate, cost credits for existing equipment or redundant equipment were also assigned. After assignment of costs for all retrofit hardware and consideration of all pertinent cost credits, capital costs, O&M costs and total annual costs were calculated for each retrofit situation. O&M costs and total annual costs were also expressed relative to the mass of phosphorus removed ($ per kg total phosphorus removed). Any license fees associated with the biological phosphorus removal processes were not included in the cost analysis.

RESULTS OF COST ANALYSIS

Retrofit Situations
The costs (capital, O&M and total annual) associated with each process retrofit to achieve phosphorus removal to each effluent phosphorus limit (1.0 mg/L and 0.3 mg/L) in each type of plant and at each selected flow are summarized in Table 3 (primary plants), Table 4 (conventional activated sludge plants) and Table 5 (extended aeration plants).

Costs of Chemical Phosphorus Removal. The retrofit of conventional chemical precipitation technology to the three plant types (primary, conventional activated sludge and extended aeration) represented the baseline case against which the biological phosphorus removal technologies were compared. Ferric chloride was used as the precipitant, at a cost of $3.70/kg Fe^{+3}.

As summarized in Table 3, the capital costs associated with retrofitting primary plants ranged from $126,000 to $159,600 to achieve a 1.0 mg/L total phosphorus limit over the size range investigated. Total annual costs ranged from $5.95/kg TP removed to $12.01/kg TP removed. For the conventional activated sludge and extended aeration plants, capital costs to achieve the 1.0 mg/L limit were comparable to the retrofit costs estimated for the primary plants. Total annual costs for the extended aeration plants were considerably higher than for the other two plant types, ranging from $8.20/kg TP removed to $18.24/kg TP removed.

Capital costs for retrofitting the secondary treatment plants to achieve the 0.3 mg/L TP limit were significantly higher. The major cost component of these retrofits was the granular filter required to achieve the lower effluent limit. In the case of conventional activated sludge plants, the total annual costs ranged from $14.46/kg TP removed to $28.03/kg TP removed over the size range investigated. Total annual costs to achieve the same effluent limit in the extended aeration plants ranged from $24.44/kg TP removed to $57.14/kg TP removed.

TABLE 3. UMMARY OF COSTS ASSOCIATED WITH RETROFITTING PRIMARY PLANTS FOR PHOSPHORUS REMOVAL

PROCESS	COST ITEM	UNITS	EFFLUENT LIMIT = 1.0 mg/L			EFFLUENT LIMIT = 0.3 mg/L		
			2.3 Ml/d	9.1 Ml/d	36.4 Ml/d	2.3 Ml/d	9.1 Ml/d	36.4 Ml/d
Chemical	Capital	$	126 000	132 000	159 600	N.A.	N.A.	N.A.
	O&M	$/yr	22 900	76 500	294 300	N.A.	N.A.	N.A.
		$/kg TP$_r$	6.91	5.76	5.54	N.A.	N.A.	N.A.
	Total Annual	$/yr	39 800	94 100	315 600	N.A.	N.A.	N.A.
		$/kg TP$_r$	12.01	7.09	5.95	N.A.	N.A.	N.A.
Bardenpho	Capital	$	1 608 600	3 124 600	7 706 400	1 759 800	3 283 000	7 987 900
	O&M	$/yr	101 500	194 000	433 000	113 000	239 700	617 600
		$/kg TP$_r$	30.63	14.62	8.16	29.02	15.37	9.90
	Total Annual	$/yr	316 900	612 300	1 464 700	348 600	679 200	1 675 000
		$/kg TP$_r$	95.61	46.14	27.59	89.52	43.56	26.85
UCT	Capital	$	1 568 800	2 943 400	7 066 800	1 720 000	3 101 800	7 258 300
	O&M	$/yr	112 000	206 000	437 800	123 500	251 700	622 400
		$/kg TP$_r$	33.80	15.52	8.25	31.72	16.14	9.98
	Total Annual	$/yr	322 000	600 100	1 383 900	353 800	667 000	1 594 100
		$/kg TP$_r$	97.16	45.22	26.07	90.86	42.77	25.56
A/O	Capital	$	1 140 600	2 227 000	4 759 200	1 291 800	2 385 400	4 950 700
	O&M	$/yr	75 500	144 000.	338 000	87 000	189 700	522 600
		$/kg TP$_r$	22.78	10.85	6.37	22.34	12.16	8.38
	Total Annual	$/yr	228 200	442 200	975 200	259 900	509 100	1 185 400
		$/kg TP$_r$	68.86	33.32	18.37	66.74	32.65	19.00
Phostrip	Capital	$	1 472 100	2 840 400	5 937 600	2 114 100	4 124 400	8 841 600
	O&M	$/yr	124 000	280 700	871 000	140 500	311 700	943 000
		$/kg TP$_r$	37.42	21.15	16.41	36.08	19.99	15.12
	Total Annual	$/yr	321 100	661 000	1 666 000	423 500	863 900	2 126 700
		$/kg TP$_r$	96.89	49.81	31.38	108.76	55.40	34.09

TABLE 4. SUMMARY OF COSTS ASSOCIATED WITH RETROFITTING CONVENTIONAL ACTIVATED SLUDGE PLANTS FOR PHOSPHORUS REMOVAL

PROCESS	COST ITEM	UNITS	EFFLUENT LIMIT = 1.0 mg/L			EFFLUENT LIMIT = 0.3 mg/L		
			4.5 Ml/d	13.6 Ml/d	36.4 Ml/d	4.5 Ml/d	13.6 Ml/d	36.4 Ml/d
Chemical	Capital	$	129 600	1.6 200	159 600	1 012 400	1 752 900	3 058 200
	O&M	$/yr	37 400	103 900	253 400	82 800	214 600	493 300
		$/kg TP$_r$	5.64	5.22	4.77	10.63	9.18	7.90
	Total Annual	$/yr	54 800	122 100	274 800	218 300	449 200	902 700
		$/kg TP$_r$	8.27	6.14	5.17	28.03	19.22	14.46
Bardenpho	Capital	$	1 813 200	3 246 000	5 766 000	1 968 700	3 409 400	5 957 500
	O&M	$/yr	67 600	124 500	209 100	91 500	193 200	393 700
		$/kg TP$_r$	10.20	6.26	3.93	11.75	8.27	6.30
	Total Annual	$/yr	310 400	559 100	981 100	355 100	649 650	1 191 300
		$/kg TP$_r$	46.83	28.12	18.46	45.60	27.80	19.08
UCT	Capital	$	1 875 600	3 186 000	5 670 00	2 031 100	3 349 400	5 861 500
	O&M	$/yr	79 200	138 500	214 900	103 100	207 200	397 600
		$/kg TP$_r$	11.95	6.97	4.04	13.24	8.87	6.37
	Total Annual	$/yr	330 300	565 000	974 000	375 000	655 600	1 182 300
		$/kg TP$_r$	49.83	28.41	18.33	48.15	28.06	18.93
A/O	Capital	$	1 272 600	2 252 400	3 828 000	1 428 100	2 415 800	4 019 500
	O&M	$/yr	37 100	66 500	115 100	61 000	135 200	297 800
		$/kg TP$_r$	5.60	3.34	2.17	7.83	5.79	4.77
	Total Annual	$/yr	207 500	368 100	627 600	252 200	458 600	835 900
		$/kg TP$_r$	31.31	18.51	11.81	32.38	19.63	13.39
Phostrip	Capital	$	867 000	1 377 600	2 101 200	1 750 200	2 997 600	5 005 200
	O&M	$/yr	108 200	262 000	608 400	129 700	302 000	680 400
		$/kg TP$_r$	16.32	13.18	11.45	16.65	12.93	10.90
	Total Annual	$/yr	224 300	446 400	889 700	364 000	703 300	1 350 500
		$/kg TP$_r$	33.84	22.45	16.74	46.74	30.10	21.63

TABLE 5. SUMMARY OF COSTS ASSOCIATED WITH RETROFITTING EXTENDED AERATION PLANTS FOR PHOSPHORUS REMOVAL.

PROCESS	COST ITEM	UNITS	EFFLUENT LIMIT = 1.0 mg/L			EFFLUENT LIMIT = 0.3 mg/L		
			0.91 Ml/d	1.6 Ml/d	3.2 Ml/d	0.91 Ml/d	1.6 Ml/d	3.2 Ml/d
Chemical	Capital	$	124 000	125 000	125 000	514 000	600 000	745 000
	O&M	$/yr	7 700	12 300	21 600	20 400	30 700	34 500
		$/kg TP$_r$	5.79	5.26	4.62	13.07	11.18	6.28
	Total Annual	$/yr	24 300	29 000	38 300	89 200	111 000	134 200
		$/kg TP$_r$	18.28	12.41	8.20	57.14	40.44	24.44
Bardenpho	Capital	$	918 000	1 140 000	1 446 000	1 066 800	1 290 000	1 596 000
	O&M	$/yr	33 100	43 700	48 300	38 300	51 100	59 200
		$/kg TP$_r$	24.90	18.71	10.34	24.54	18.62	10.78
	Total Annual	$/yr	156 000	196 300	241 900	181 100	223 800	272 900
		$/kg TP$_r$	117.38	84.04	51.77	116.03	81.53	49.70
UCT	Capital	$	954 000	1 152 000	1 512 000	1 102 800	1 302 000	1 662 000
	O&M	$/yr	43 100	52 700	60 300	48 300	60 100	71 200
		$/kg TP$_r$	32.43	22.56	12.91	30.94	21.89	12.97
	Total Annual	$/yr	170 800	206 900	262 700	195 900	234 400	293 700
		$/kg TP$_r$	128.52	88.57	56.23	125.50	85.39	53.50
A/O	Capital	$	757 200	895 200	1 146 600	906 000	1 045 200	1 296 600
	O&M	$/yr	27 100	31 700	30 300	32 300	39 100	41 200
		$/kg TP$_r$	20.39	13.57	6.49	20.69	14.24	7.50
	Total Annual	$/yr	128 500	151 500	183 800	153 600	179 000	214 800
		$/kg TP$_r$	96.69	64.85	39.34	98.40	65.21	39.13
Phostrip	Capital	$	410 400	529 200	740 400	878 400	1 099 200	1 484 400
	O&M	$/yr	39 000	49 500	82 000	46 000	60 500	99 000
		$/kg TP$_r$	29.35	21.19	17.55	29.47	22.04	18.03
	Total Annual	$/yr	93 900	120 300	181 100	163 600	207 700	297 700
		$/kg TP$_r$	70.65	51.50	38.76	104.80	75.66	54.22

Costs of Bardenpho Process. Retrofitting the model plants to achieve phos-phorus removal by the Bardenpho process was considerably more capital cost intensive than the retrofit of chemical phosphorus removal processes, in all cases. However, the capital costs associated with achieving an effluent limit of 0.3 mg/L were not significantly higher than the capital costs asso-ciated with achieving an effluent limit of 1.0 mg/L. An effluent filter must be incorporated into both flowsheets. The only additional equipment required to achieve the lower effluent limit was the chemical feed equipment asso-ciated with ferric chloride dosing to the process.

Figures 8 and 9 compare the total annual costs of the Bardenpho retrofit to the costs of chemical phosphorus removal for the three plant types evaluated to effluent limits of 1.0 mg/L and 0.3 mg/L, respectively. It is apparent that unit costs ($/kg TP removed) increased dramatically as plant sizes decreased. For the smallest plants, total annual costs were in the order of $100/kg TP removed. For all plant sizes evaluated, chemical precipitation technology was less costly than the retrofit of the Bardenpho process. The most cost competitive situation was the retrofit of the largest conventional activated sludge plant (36.4 Ml/d) to the lower phosphorus limit. In this case, the total annual costs of the Bardenpho process were approximately 30 percent higher than those associated with the chemical phos-phorus removal process.

Costs of UCT Process. The costs of retrofitting the UCT process were compar-able to the costs of the Bardenpho process retrofit in all of the cases eval-uated. Total annual costs were not particularly sensitive to the effluent phosphorus limit imposed on the process, but were very sensitive to plant size. On a unit cost basis ($ per kg TP removed), retrofitting of conven-tional activated sludge plants represented the most inexpensive situation.

Figures 10 and 11 present the total annual costs associated with the UCT retrofit situations for the two effluent total phosphorus limits imposed. As was the case for the Bardenpho process, chemical retrofit was the more cost effective alternative in all situations. It is interesting to note, however, that the annual O&M costs of the biological system were less than the O&M costs of the chemical system for the larger conventional activated sludge plants (Table 4). The total annual costs were higher due to the high capital costs associated with the UCT retrofit.

Costs of A/O Process. The capital costs associated with retrofitting the A/O process for biological phosphorus removal were considerably lower than the capital costs of the Bardenpho and UCT process retrofits in all cases. The lower costs related to the reduced equipment requirements needed for the more simplistic two-stage A/O process compared to the other two biological proces-ses. A/O capital costs were also relatively insensitive to the effluent phosphorus limit imposed.
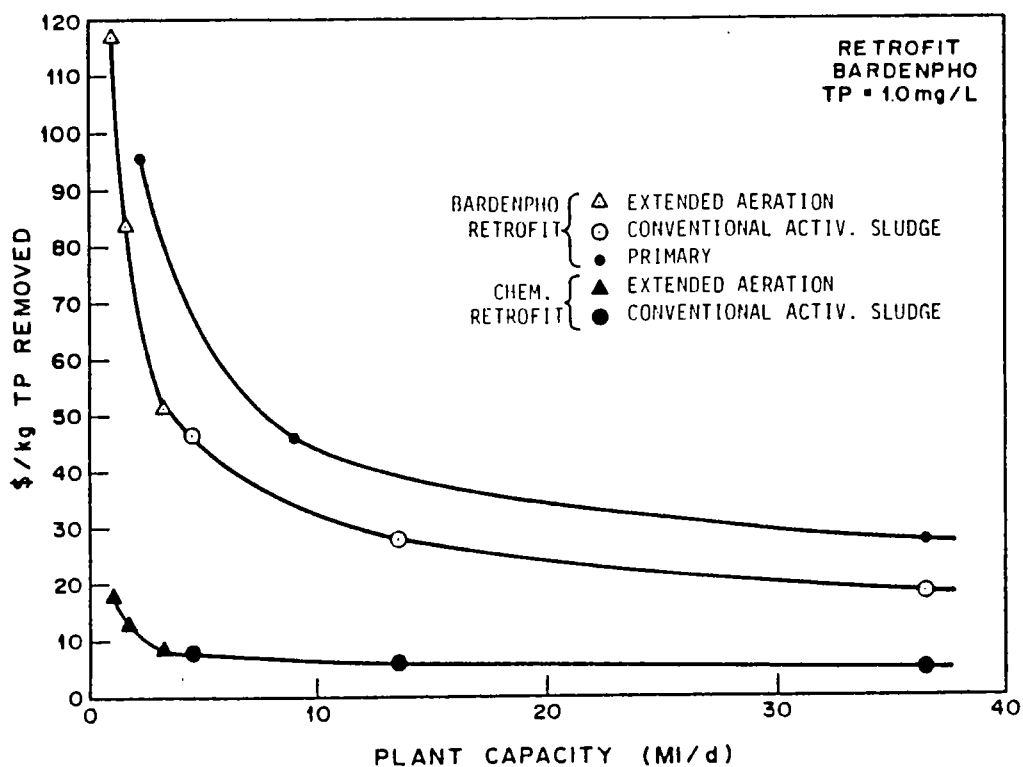
FIGURE 8: RETROFIT TOTAL ANNUAL COSTS – BARDENPHO TECHNOLOGY
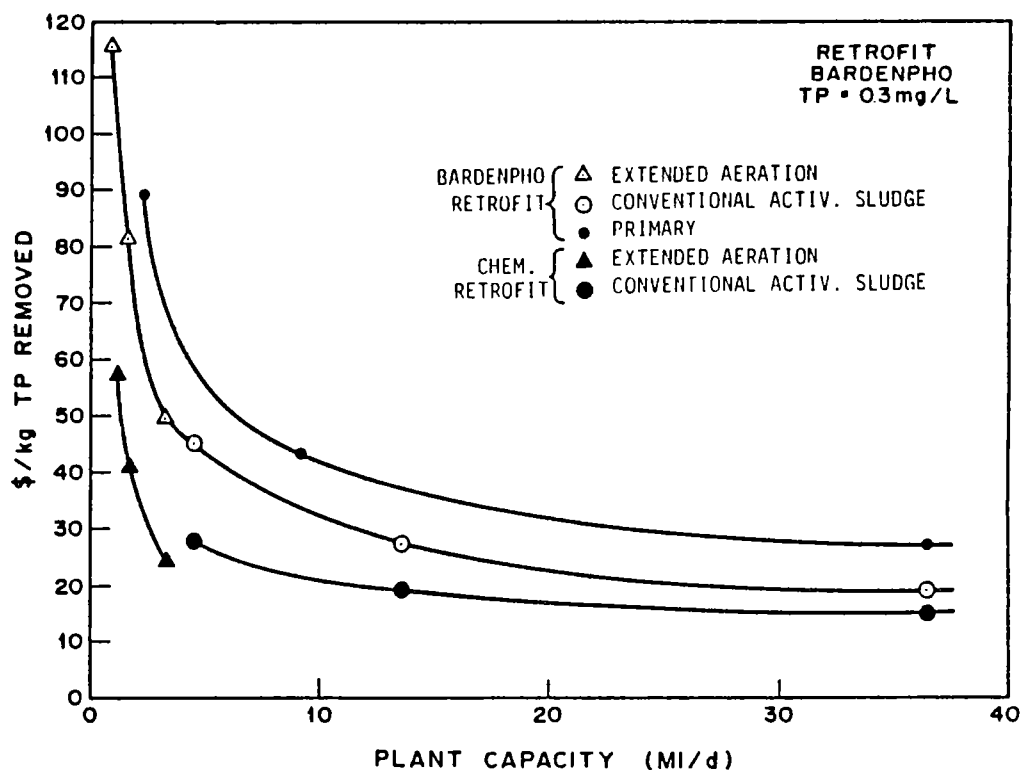TO 1.0 mg/L TP

FIGURE 9: RETROFIT TOTAL ANNUAL COSTS – BARDENPHO TECHNOLOGY
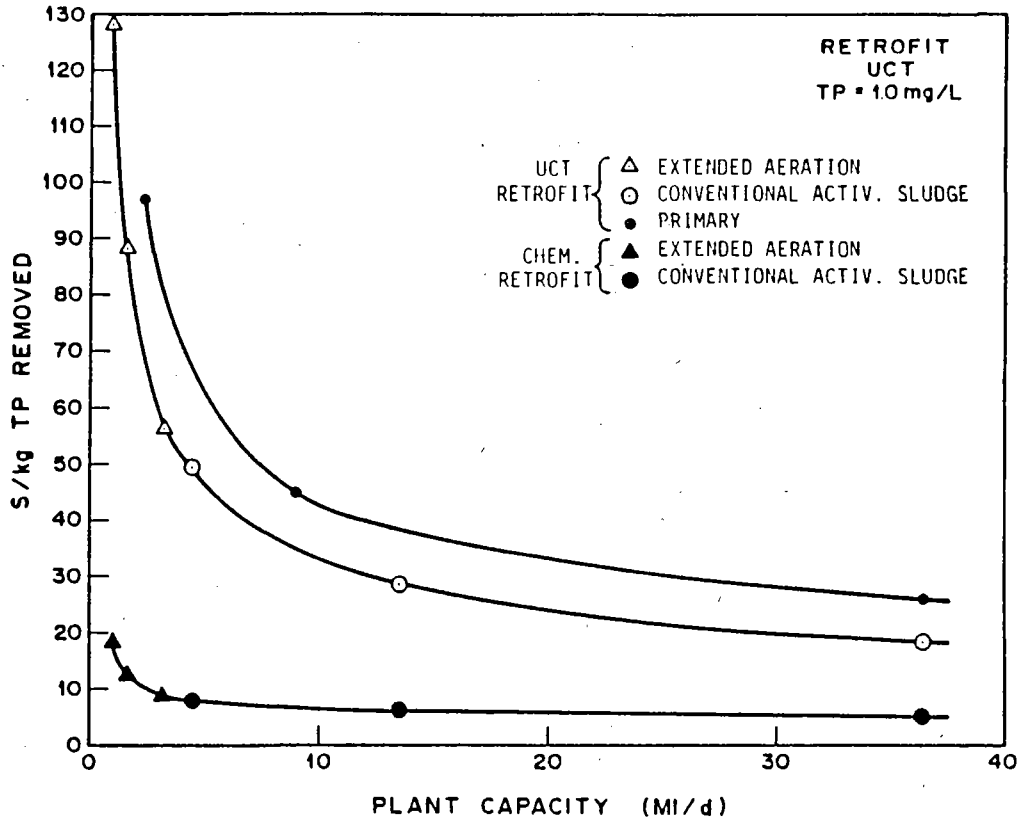TO 0.3 mg/L TP

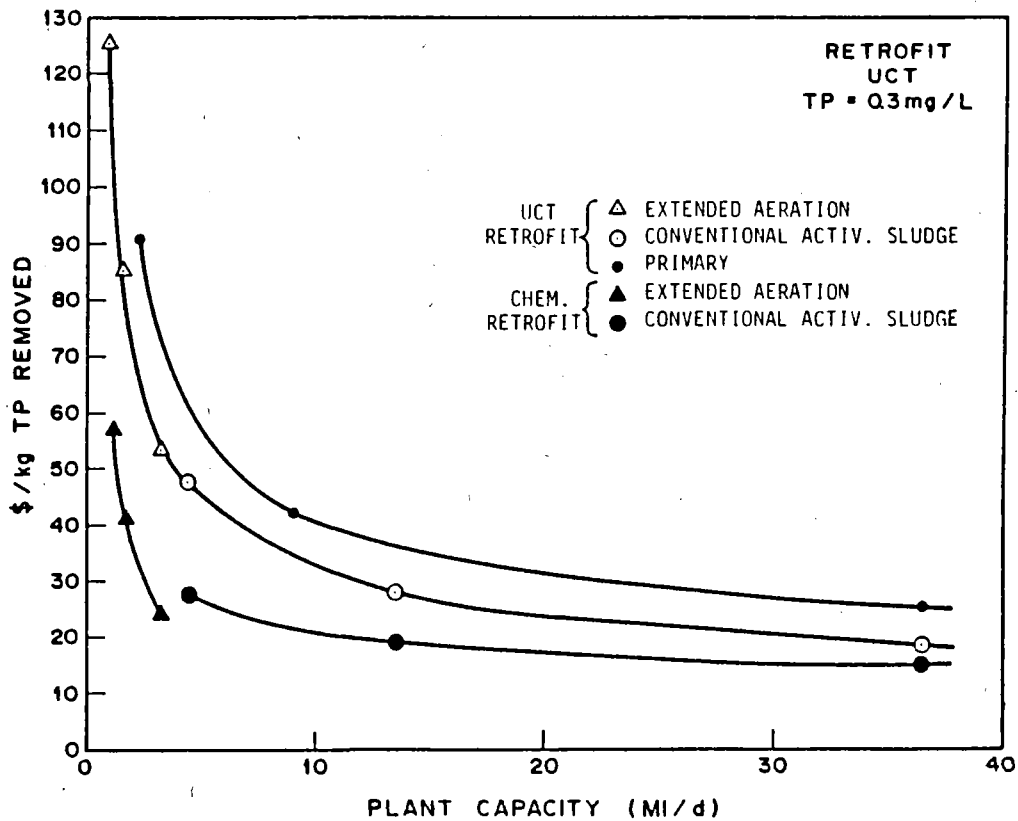FIGURE 10: RETROFIT TOTAL ANNUAL COSTS – UCT TECHNOLOGY
TO 1.0 mg/L TP

FIGURE 11: RETROFIT TOTAL ANNUAL COSTS – UCT TECHNOLOGY
TO 0.3 mg/L TP

The total annual costs of the A/O retrofits are presented in Figure 12 (1.0 mg/L effluent TP limit) and Figure 13 (0.3 mg/L effluent TP limit). At the higher effluent limit, retrofit of chemical precipitation technology is more cost effective in all cases. However, at the lower effluent limit, the A/O process was cost competitive with the chemical precipitation technology in conventional activated sludge plants treating flows in excess of approximately 20 Ml/d. At a flow of 36.4 Ml/d, the largest plant size evaluated, the total annual costs associated with the A/O process were approximately 7 percent lower than comparative costs for the chemical precipitation process. Annual O&M costs of the A/O process were lower than the O&M costs of the chemical phosphorus removal process for all conventional activated sludge plant retrofit situations. It should be noted that licensing fees for the A/O process have not been included in these costs. At the highest flow, the licensing fee would add approximately $0.90/kg TP removed to the total annual costs of the A/O process.

Costs of Phostrip Process. Capital costs of retrofitting the Phostrip process into the secondary plants (conventional activated sludge and extended aeration) to achieve the higher effluent TP limit were lower than the comparative costs of retrofitting the other three biological process options. However, the Phostrip retrofit was considerably more sensitive to the effluent limit imposed and capital costs of achieving the lower TP limit by the Phostrip process were comparable to costs of achieving the same effluent limit by the other processes. The substantial increase in capital costs of the Phostrip retrofit to achieve the lower effluent limit were associated with the effluent filter which was not required to achieve a limit of 1.0 mg/L using Phostrip.

Annual O&M costs of the Phostrip retrofit were higher than the O&M costs of any of the other biological options in all conventional activated sludge retrofit situations. Figures 14 and 15 present the total annual costs ($ per kg TP removed) for the Phostrip retrofits. In all cases, retrofit of conventional chemical precipitation technology was more cost effective than retrofit of Phostrip technology.

New Plant Costs

Capital, O&M and total annual costs were calculated for new facilities achieving the selected effluent phosphorus targets of 1.0 mg/L and 0.3 mg/L. The costs of the four biological process options (Bardenpho, UCT, A/O and Phostrip) were compared with the costs of a conventional activated sludge plant utilizing chemical phosphorus precipitation. The cost analysis was done for plant sizes of 4.5 Ml/d, 13.6 Ml/d and 36.4 Ml/d.

Total annual costs ($ per kg TP removed) are presented in Figures 16 and 17 for effluent TP limits of 1.0 mg/L and 0.3 mg/L, respectively. The capital costs of the biological phosphorus removal options were generally higher than the capital costs of the chemical phosphorus removal options. Only the A/O process, designed to achieve an effluent limit of 0.3 mg/L, resulted in lower capital costs than the chemical option. Annual O&M costs of the chemical and biological options were generally competitive. However, as shown in Figure 16, none of the biological options were competitive with the conventional chemical precipitation process to an effluent limit of 1.0 mg/L in terms of the total annual costs. To an effluent limit of 0.3 mg/L, the A/O process was most cost effective at all plant sizes. Furthermore, the costs of all process alternatives were within ± 10 percent of the costs of the chemical precipitation alternative.
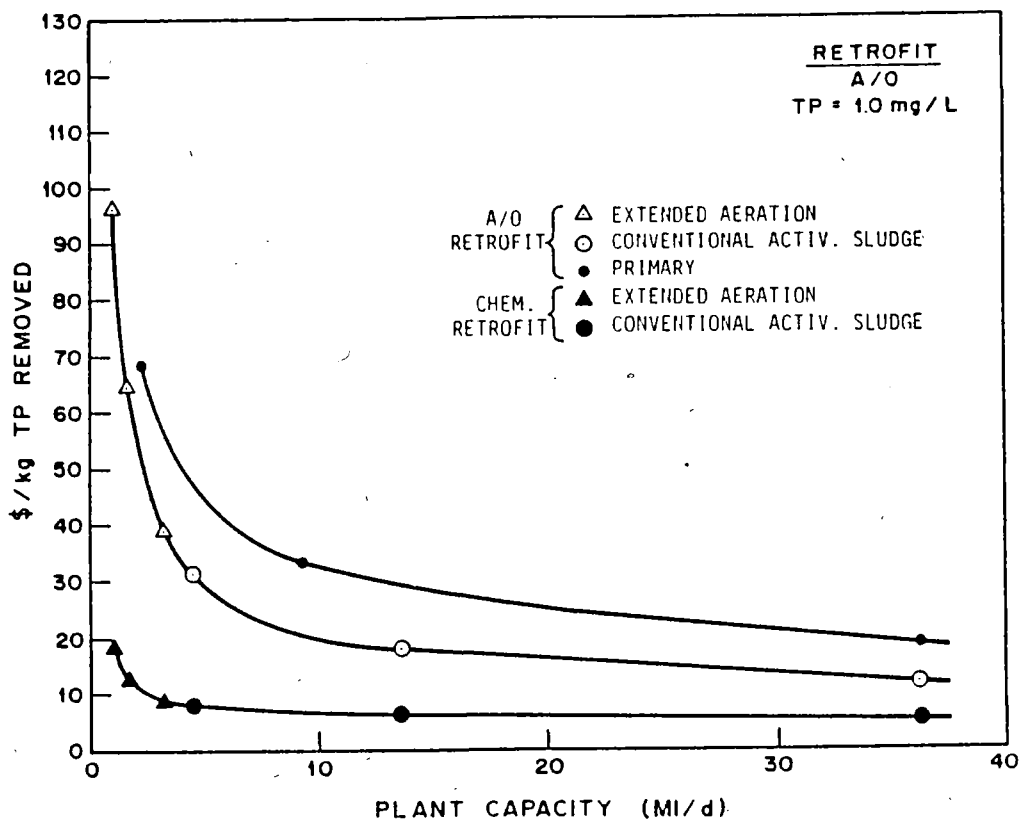
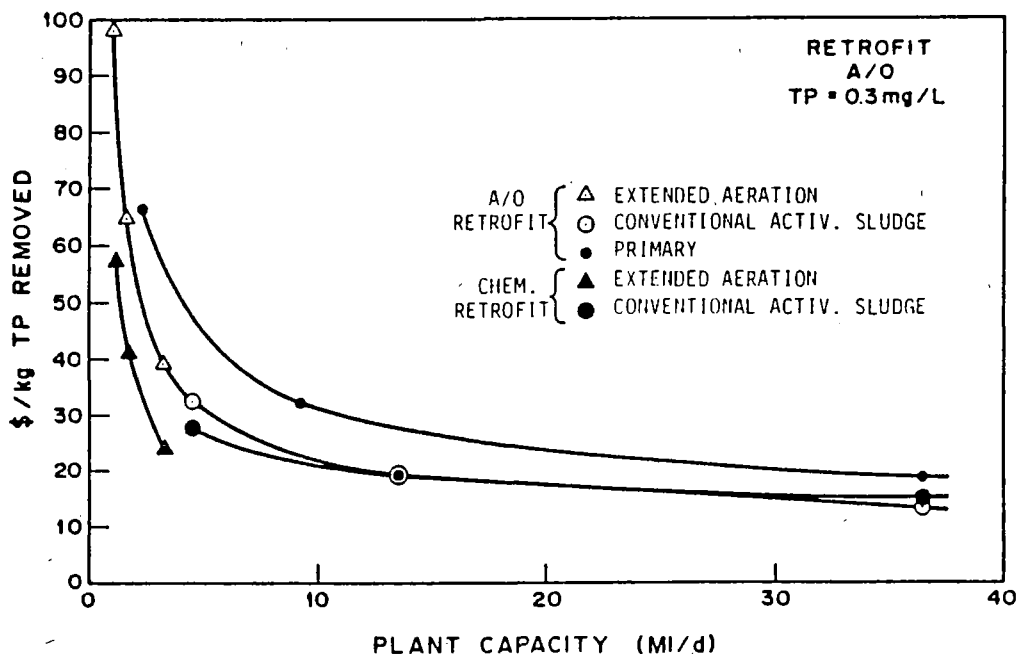**FIGURE 12: RETROFIT TOTAL ANNUAL COSTS – A/O TECHNOLOGY TO 1.0 mg/L TP**

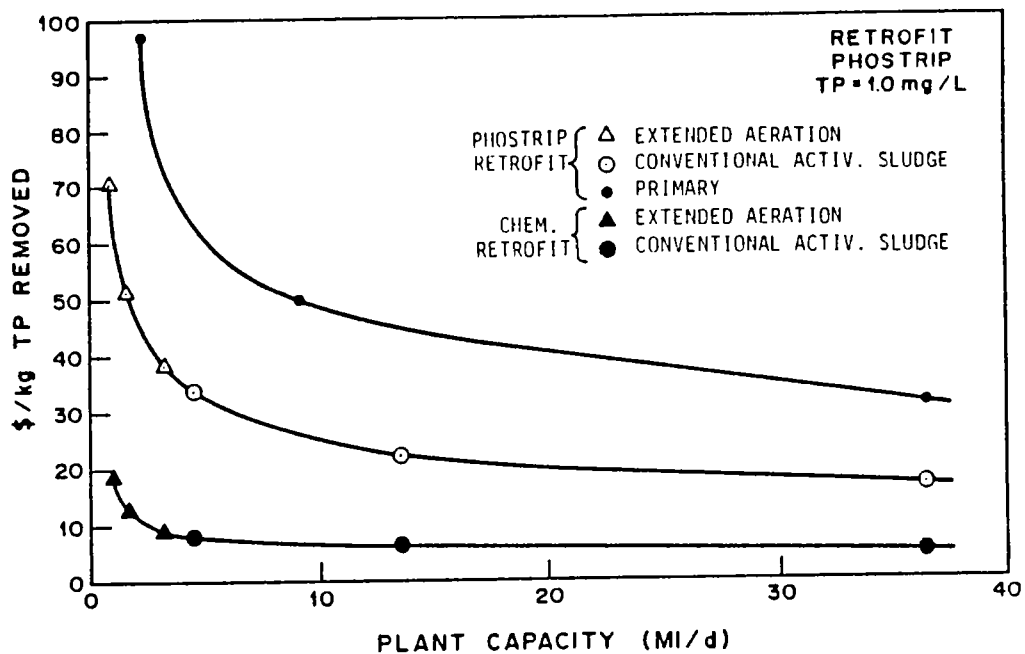**FIGURE 13: RETROFIT TOTAL ANNUAL COSTS – A/O TECHNOLOGY TO 0.3 mg/L TP**

FIGURE 14: RETROFIT TOTAL ANNUAL COSTS – PHOSTRIP TECHNOLOGY
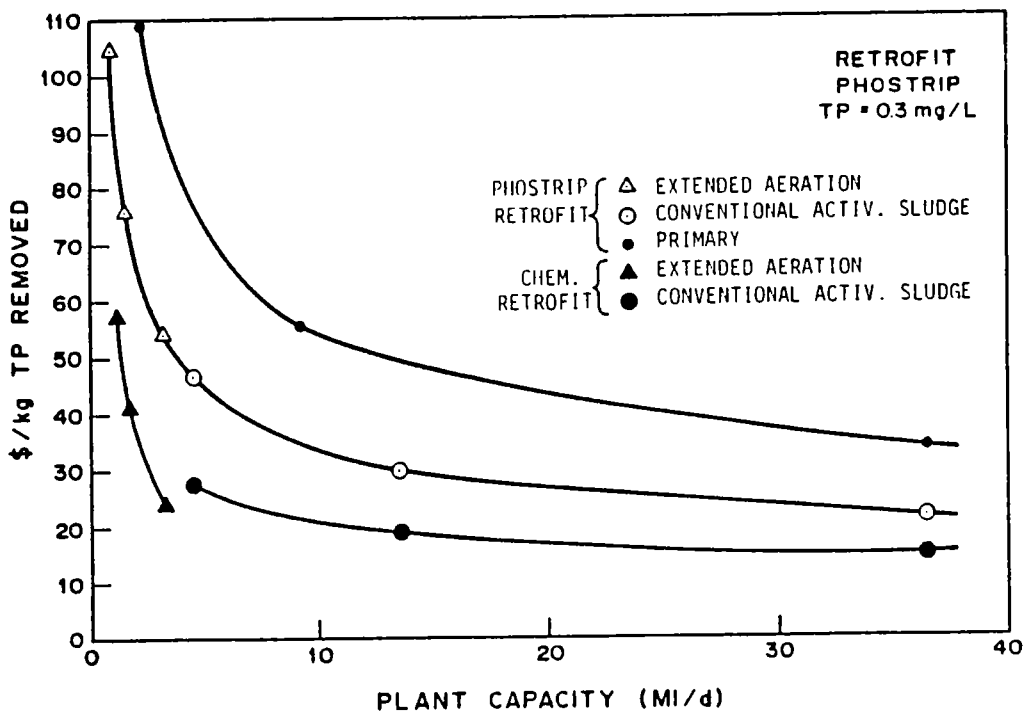TO 1.0 mg/L TP

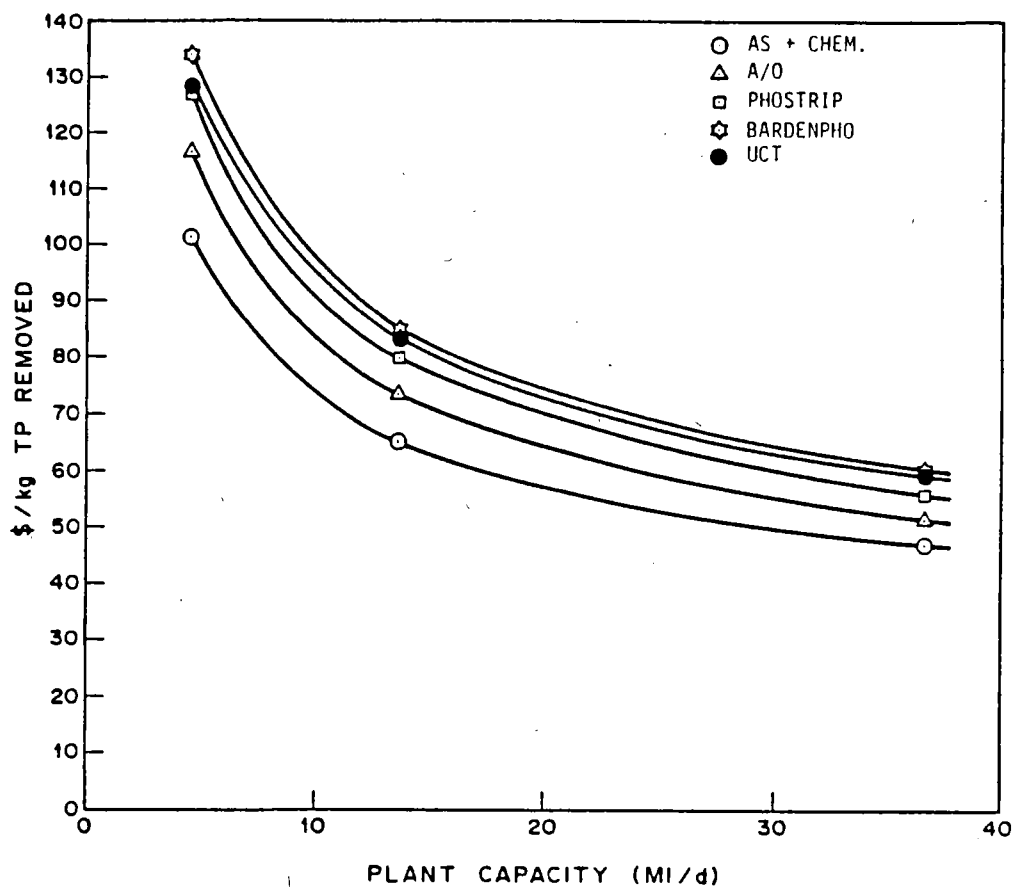FIGURE 15: RETROFIT TOTAL ANNUAL COSTS – PHOSTRIP TECHNOLOGY
TO 0.3 mg/L TP

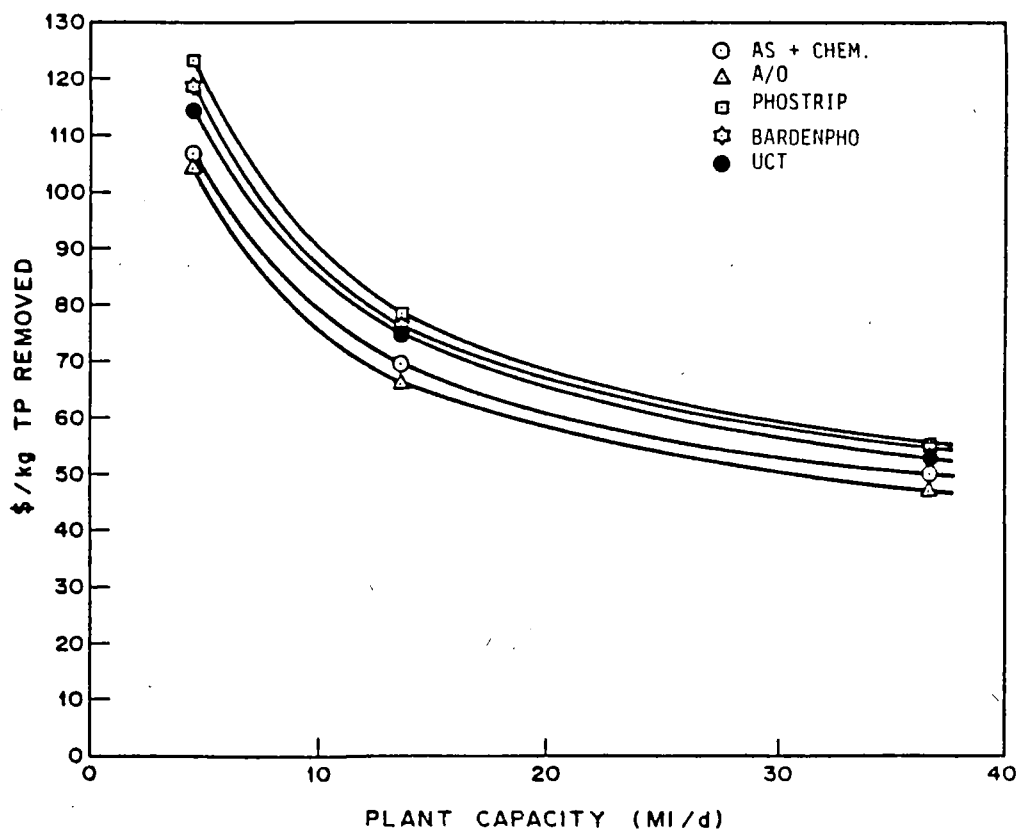FIGURE 16: TOTAL ANNUAL COSTS EBPR TO 1.0 mg/L – NEW PLANTS

FIGURE 17: TOTAL ANNUAL COSTS EBPR TO 0.3 mg/L – NEW PLANTS

Sensitivity of Chemical Precipitation Process Costs

The total annual costs of the conventional chemical precipitation technology were found to be relatively insensitive to increases in chemical dosage and in chemical costs. As shown in Figure 18, increasing the ferric chloride dosage by a factor of two resulted in an increase in the total annual costs of between 4 and 12 percent depending on the effluent TP limit and plant flow. Increasing the cost of ferric chloride by a factor of two resulted in an increase in the total annual costs of between 2 and 5 percent. The greater sensitivity to dosage related to the increase in sludge disposal costs produced at higher chemical addition rates.
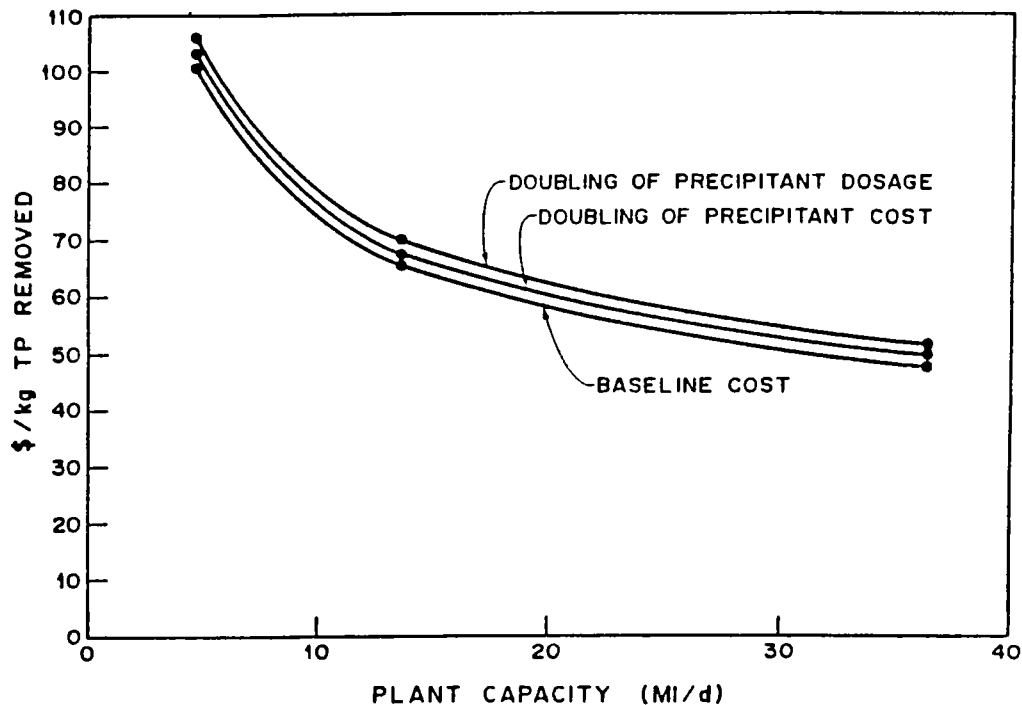
FIGURE 18: COST SENSITIVITY OF CHEMICAL PRECIPITANT
- TOTAL ANNUAL COST - NEW PLANTS

Substituting alum for ferric chloride increased the total annual costs of the chemical precipitation option by between 3 and 6 percent depending on the effluent TP limit and plant size.

SUMMARY AND CONCLUSIONS

Based on the technical review of biological phosphorus removal processes, the following conclusions can be drawn:

- To consistently achieve an effluent total phosphorus limit of 1.0 mg/L by the application of either the Bardenpho, UCT or A/O processes will require final effluent filtration. The Phostrip process will consistently achieve this effluent limit without effluent filtration.

- To achieve an effluent total phosphorus limit of 0.3 mg/L, the Bardenpho, UCT and A/O processes must be supplemented by chemical addition and effluent filtration. The Phostrip process required only effluent filtration to achieve this effluent total phosphorus limit.

Based on the cost analyses conducted, the following summarizes the most pertinent findings:

- For plant sizes up to 36.4 Ml/d and an effluent target of 1.0 mg/L total phosphorus, it is more cost effective to retrofit existing extended aeration, conventional activated sludge and primary plants for chemical precipitation of phosphorus. Chemical precipitation technology is also more cost effective in new plants under these conditions.
- For an effluent target of 0.3 mg/L, it is more cost-effective to retrofit extended aeration plants and conventional activated sludge plants with chemical precipitation technology for flows up to approximately 4.5 Ml/d. The A/O process becomes more cost-effective for retrofitting conventional activated sludge plants at flows in excess of approximately 13.6 Ml/d.
- In new plants, the A/O process is more cost effective than chemical precipitation technology at flows as low as approximately 3.2 Ml/d.

REFERENCES

1. Ontario Ministry of the Environment, "Guidelines for the Design of Sewage Treatment Works", Toronto, Ontario, (1984).

2. Harris, R.W., M.J. Cullinane, Jr., and P.T. Sun, "Process Design and Cost Estimating Algorithms for the Computer Assisted Procedure for Design and Evaluation of Wastewater Systems (CAPDET)", prepared for the U.S. Environmental Protection Agency by the Environmental Engineering Division of the U.S. Army Engineer Waterways Experiment Station, Vicksburg, Mississippi, (1982).

3. Pineau, M., P. Cote, and J.P. Villeneuve, "Estimation of Wastewater Treatment Costs : Evaluation of the CAPDET Model for Canadian Conditions", Accepted for publication, Canadian Journal of Civil Engineering, (1985).

4. Paepcke, B.H., "Introduction to Biological Phosphorus Removal", presented at the Technology Transfer Seminar on Biological Phosphorus Removal in Municipal Wastewater Treatment, Penticton, British Columbia, (April 1985).

5. Canviro Consultants Ltd., "Technical and Economic Feasibility of Retrofitting Existing Municipal Wastewater Treatment Plants in Canada for Biological Phosphorus Removal", draft report to the Environmental Protection Service, Environment Canada, (March 5, 1985).

6. Marais, G.v.R. and G.A. Ekama, "The Activated Sludge Process - Part I - Steady State Behaviour", Selected Papers on Activated Sludge Process Research at the University of Cape Town, Cape Town, R.S.A., (1982).

7. Prested, B.P., E.E. Shannon and R.J. Rush, "Development of Prediction Models for Chemical Phosphorus Removal, Volume 1", Canada-Ontario Agreement Research Report No. 68, Environmental Protection Service, Environment Canada, Ottawa, Ontario, (1977).

8. U.S. Environmental Protection Agency, "Process Design Manual for Phosphorus Removal", EPA 625/1-76-001a, Washington, D.C., (1976).

9. Schmidtke, N.W., "Sludge Generation, Handling and Disposal at Phosphorus Control Facilities in Ontario", in Characterization, Treatment and Use of Sewage Sludge, Proceedings of the Second European Symposium, Vienna, Austria, (1980).

RETROFIT OF A FULL MUNICIPAL WASTEWATER

TREATMENT PLANT AT PONTIAC, MICHIGAN

FOR BIOLOGICAL PHOSPHORUS REMOVAL

S.J. Kang

McNamee Porter & Seeley

Ann Arbor, Michigan

RETROFIT OF A FULL SCALE
MUNICIPAL TREATMENT PLANT AT PONTIAC, MICHIGAN
FOR BIOLOGICAL PHOSPHORUS REMOVAL

Shin Joh Kang,
McNamee, Porter and Seeley, Ann Arbor, Michigan
and
Paul J. Horvatin,
U.S. Environmental Protection Agency, Chicago, Illinois

INTRODUCTION

The City of Pontiac, Michigan operates two wastewater treatment plants: the East Boulevard Plant with secondary treatment using the activated sludge process, and the Auburn Plant, also using the activated sludge process with tertiary sand filters. Secondary effluent from the East Boulevard Plant is treated by Auburn's sand filters prior to discharge. The combined effluent from the Auburn Plant must meet the following NPDES (National Pollutant Discharge Elimination System) Permit effluent limits shown in Table 1.

TABLE 1
NPDES PERMIT EFFLUENT LIMITS

PARAMETER	DATES	DAILY MAX. MG/L	30-DAY AVG. MG/L	7-DAY AVG. MG/L
Carbonaceous BOD$_5$	May 1 to Nov 30	10	7	--
	Dec 1 to Mar 31	22	15	--
	All Year	--	30	45
Ammonia Nitrogen	May 1 to Sept 30	3.2	--	--
	Oct 1 to Nov 30	--	6.7	--
	Dec 1 to Mar 31	--	9.0	--
	Apr 1 to Apr 30	--	10.3	--
Total Suspended Solids	May 1 to Nov 30	--	20	30
	Dec 1 to Apr 30	--	30	45
Total Phosphorus	All Year	--	1.0	--

The current method of phosphorus removal at the Auburn Plant and the method used at the East Boulevard Plant prior to this demonstration project involve the use of iron salts and polymer for chemical precipitation of phosphorus in the primary clarifiers. Chemicals used for this precipitation are costly, and a good deal of chemical sludge is produced as a result, which adds to the overall cost of sludge disposal.

The A/O biological phosphorus removal process is similar to the conventional activated sludge process, except that primary effluent and return activated sludge undergo a short anaerobic detention prior to entering the

aeration basin. The continuous anaerobic/aerobic cycling of the activated sludge promotes the growth of microorganisms capable of assimilating increased levels of phosphorus in the aeration basin over that of conventional activated sludge. The A/O process has been successfully operated at a full-scale facility in the warm climate of Largo, Florida; however, full-scale operation in a cold climate had not been demonstrated. Since most wastewater treatment facilities in the Great Lakes region are required to meet stringent effluent phosphorus limits, the successful demonstration of this biological phosphorus removal process would provide an attractive alternative to the more conventional chemical precipitation processes.

In addition to demonstration of the A/O process in a cold climate, several other factors required investigation. Some wastewater treatment facilities in this region are also required to produce effluents low in ammonia nitrogen during the summer months, and a few are required to do so in the winter months. The nitrification process used to achieve low effluent ammonia concentrations also produces high levels of nitrate nitrogen in the effluent and therefore in the return activated sludge. Nitrate recycled in the return sludge to the anaerobic stages was claimed to be detrimental to the process by some researchers. Thus, operation of the A/O process in conjunction with the nitrification process required investigation.

Another complicating factor for the A/O process was the widespread use of anaerobic digestion for sludge processing. The supernatant from anaerobic digesters is typically recycled through the primary and secondary processes. Under anaerobic conditions, however, a portion of the phosphorus assimilated by the microorganisms under aeration is released back into a soluble form. This soluble phosphorus would then be recycled along with the digester supernatant to the secondary biological process. The effect of this increase in phosphorus load as well as any other detrimental effects of the supernatant recycle needed further investigation.

After initial review of Pontiac's wastewater treatment facilities configurations and operations, it was determined that Pontiac would be an ideal community for demonstration of the A/O biological phosphorus removal process. The East Boulevard Plant consists of four identical activated sludge trains, two of which could easily be converted to the A/O process, leaving two trains for comparison. In addition, liquid and sludge streams between the East Boulevard and Auburn Plants were connected by piping. Thus, during periods without digester supernatant recycle, the supernatant could be treated at the Auburn Plant. In addition, if process upsets occurred at the East Boulevard Plant, the effluent would still undergo tertiary treatment at the Auburn Plant, ensuring that NPDES Permit effluent limits would be met.

RETROFIT

PLANT LAYOUT

A configuration of the East Boulevard Plant as it existed prior to retrofit of one-half to the A/O process is shown in Figure 1. Four identical plug-flow aeration trains provided secondary treatment.

Figure 2 shows the plant configuration after retrofit. Each treatment train is 109 m (358 feet) long and 6.1 m (20 feet) wide with 3.35 m (11 feet) side water depth. Two plastic baffles suspended under a plastic collar were used to delineate three anaerobic stages in each train approximately 7.6 m
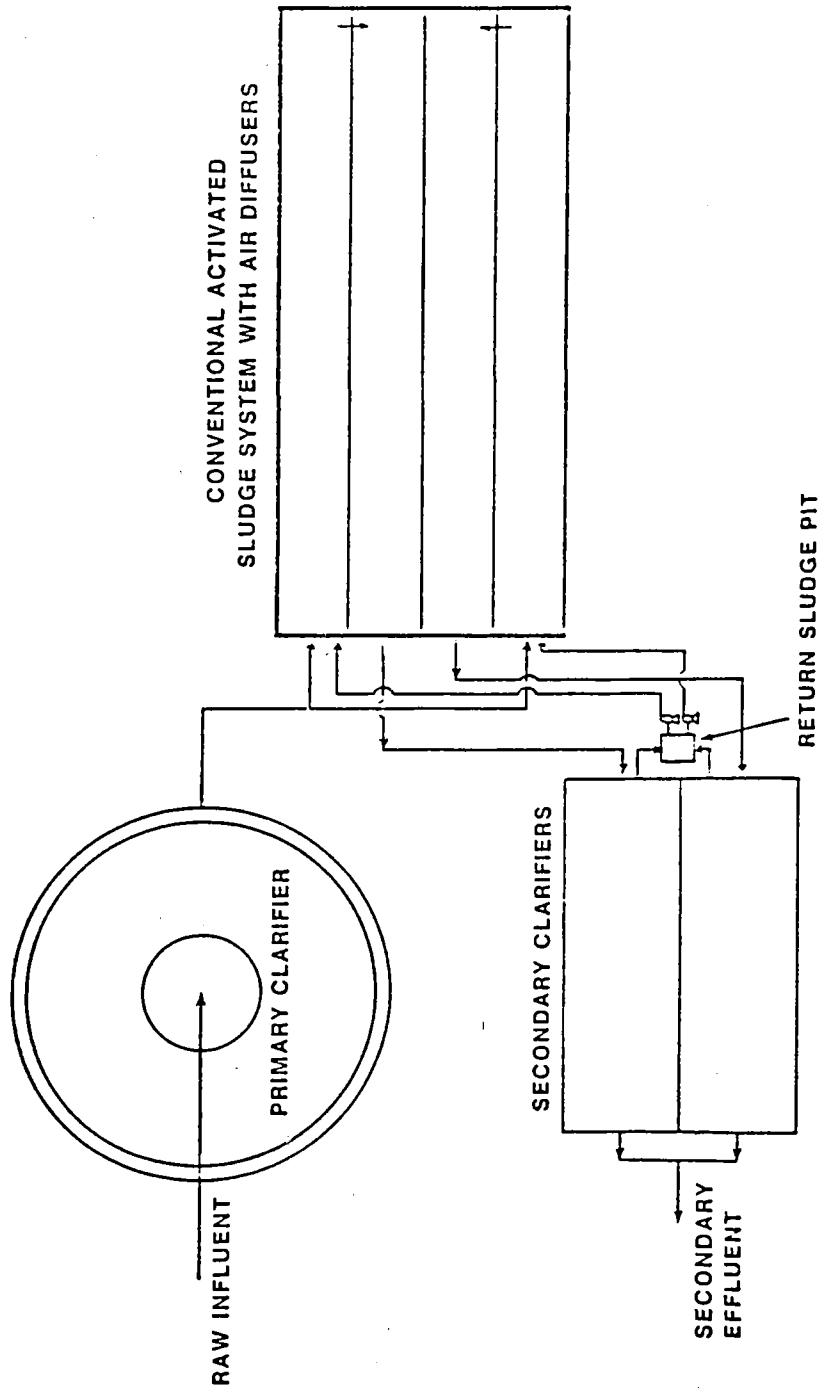
FIGURE 1

EXISTING BASIN CONFIGURATION OF
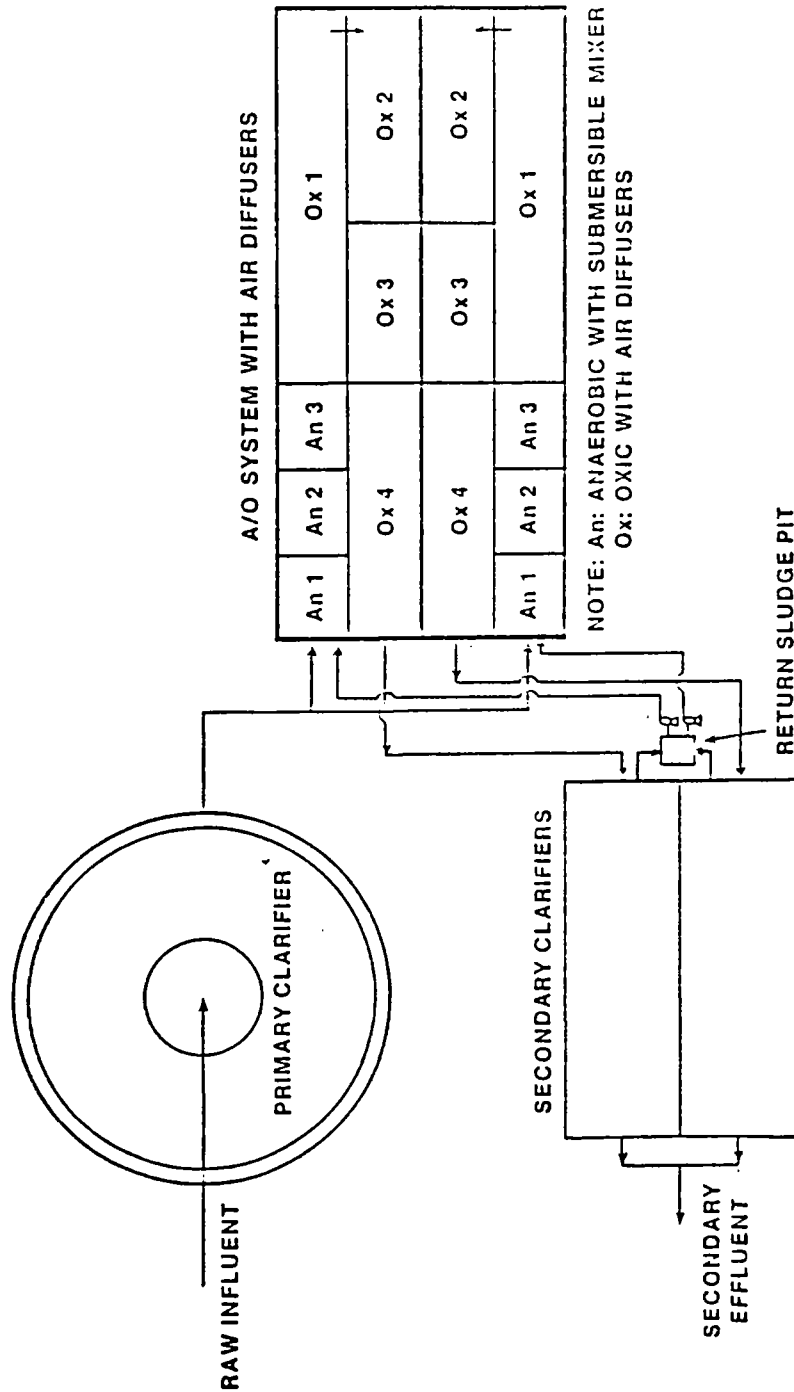CONVENTIONAL ACTIVATED SLUDGE SYSTEM

FIGURE 2

RETROFITTED A/O SYSTEM FOR
BIOLOGICAL REMOVAL

(25 feet) in length. The remaining length in the first half of each train was used as the first aerobic, or oxic stage. The second, third, and fourth oxic stages in the second half of each train were separated by wooden baffles constructed of Wolmanized southern pine with lengths of 18.3, 13.6 and 23.1 m (60, 44.5 and 76 feet), respectively. One wooden baffle also separated the last anaerobic and first oxic stage. Three foot square openings located at the bottom of each baffle on alternating sides allowed movement of mixed liquor from one stage to the next. In addition, small slots at the top of the baffles allowed movement of scum through each train.

A diffused air system provided aeration for the activated sludge. Diffuser socks in the anaerobic stages were removed and the existing lines were plugged. Side-mounted submersible Flygt mixers were installed in each stage to provide mixing for the biological solids. The existing diffused air system provided aeration for the oxic stages.

The installation of baffles and mixers were the only alterations required for conversion to the A/O process.

SCHEDULE AND COSTS

After acceptance of a U.S. EPA Demonstration Grant by the City of Pontiac on October 11, 1983, a project coordination meeting was held on October 14, 1983 among the City, APCI (Air Products and Chemicals, Inc.) who hold the patent for the A/O process, and the City's engineers, the consulting firm of McNamee, Porter and Seeley. A design package for the A/O retrofit was submitted to Pontiac by APCI on November 4, 1983 and the City issued purchase orders for the materials necessary for the retrofit. All of the necessary materials had arrived in Pontiac by January 25, 1984 and Train 2 was dewatered for conversion. By February 10, continuous flow through the newly retrofitted A/O Train 2 had begun. Continuous flow through retrofitted A/O Train 1 began February 29. Trains 3 and 4 remained as conventional activated sludge trains for comparison. On February 7, the use of iron salts for chemical phosphorus removal was discontinued.

The cost of six 5.0 Hp Flygt mixers used to retrofit two trains was $30,000. Field testing has been recently completed to determine the minimum horsepower required for mixing the anaerobic stages. The cost of the plastic and wooden baffles for both trains was approximately $20,000. Some of the retrofit work was performed by the City's personnel.

PERFORMANCE

Evaluation of the A/O process consisted of four phases:

I   Nitrification without anaerobic digester supernatant recycle
II   Nitrification with digester supernatant recycle
III No nitrification without digester supernatant recycle
IV   No nitrification with digester supernatant recycle

The process was initially started up with a target constant flow of 7570 cu meters per day (2.0 mgd) per train. On May 11, the average flow per train was decreased to 6435 cu. meters per day (1.7 mgd) and MLSS was increased to between 2500 and 3000 mg/l to begin nitrification. Beginning June 18, a diurnal flow schedule was followed by manual adjustment of primary effluent

flow rate accomplished by plant operators. The maximum and minimum flows maintained under this schedule were 9460 and 3400 cu meters per day (2.5 and 0.9 mgd), respectively. Since nitrification was not consistently achieved under this flow schedule due to oxygen limitation, the diurnal flows were decreased to an average of 5300 cu meters per day (1.4 mgd) beginning July 13.

In the normal mode of operation at the East Boulevard Plant, waste activated sludge is returned to the primary clarifiers for co-settling with raw sludge. In an effort to clear the anaerobic digesters from sludge containing iron, which could affect process performance during periods of digester supernatant recycle, only raw sludge was pumped to the digesters, with waste activated sludge pumped to the Auburn Plant for processing. Waste activated sludge was once again returned to the primary clarifiers beginning August 21, with full digester supernatant recycle beginning October 1, 1984. Nitrification continued throughout this period.

On November 15, digester supernatant recycle was discontinued and primary effluent flows were returned to the previous diurnal schedule with an average flow of 6435 cu meters per day (1.7 mgd). to cease nitrification and begin Phase III. Digester supernatant was once again recycled beginning February 6, 1985 for evaluation of Phase IV.

EFFLUENT CHARACTERISTICS

Table 2 provides a summary of operation performance during the four periods of study. Averages given for Phase II exclude the month of October due to process upsets of both nitrogen and phosphorus removal encountered during this period. These upsets are thought to be a result of extreme fluctuations in raw influent pH due to an industrial discharger. Raw influent pH values were measured as low as 2 and as high as 11 pH units. The industry was contacted and the extreme pH discharges were subsequently eliminated.

With the exception of two periods of upset during mid-September and October attributed to the pH problems discussed above, the removal of phosphorus in the A/O trains was very good, and consistently higher than in the conventional trains. Average total phosphorus concentrations in the secondary effluent of the A/O trains were below 1.0 mg/l for each phase of this study.

Figure 3 gives a typical profile of total and soluble phosphorus through the A/O and conventional trains for a period with nitrification and without digester supernatant recycle. The release and uptake of soluble phosphorus in the anaerobic and oxic stages, respectively, is apparent.

Removal of ammonia was similar in the A/O and conventional trains. There were periods of nitrification upset in October and early November, probably also due to large influent pH swings. A/O secondary effluent nitrate levels were consistently lower than for the conventional secondary effluent, indicating that consistent denitrification was taking place in the A/O clarifiers.

TABLE 2
SUMMARY OF PROCESS PERFORMANCE

Parameter	PHASE I			PHASE II		
	Primary Effluent	Secondary Effluent		Primary Effluent	Secondary Effluent	
		Conven.	A/O		Conv.	A/O
Total Phosphorus, mg/l	3.24	1.92	0.77	4.12	2.26	0.68
Soluble Phosphorus, mg/l	1.86	1.75	0.69	2.23	1.95	0.59
Ammonia Nitrogen, mg/l	15.2	1.10	0.93	17.8	3.12	2.84
Nitrate Nitrogen, mg/l	0.19	15.5	10.4	0.06	13.9	11.6
Total BOD_5, mg/l	110	6.8	6.2	137	11.9	9.4
Soluble BOD_5, mg/l	65	2.5	1.8	64	2.7	3.0
Wastewater Temperature ^{o}C (^{o}F)	17 (63)			16 (61)		

Parameter	PHASE III			PHASE IV		
	Primary Effluent	Secondary Effluent		Primary Effluent	Secondary Effluent	
		Conven.	A/O		Conv.	A/O
Total Phosphorus, mg/l	3.74	1.74	0.44	2.95	1.45	0.65
Soluble Phosphorus, mg/l	2.18	1.27	0.28	1.57	1.19	0.49
Ammonia Nitrogen, mg/l	16.1	2.06	5.93	18.5	2.42	4.54
Nitrate Nitrogen, mg/l	0.20	12.9	6.65	--	14.8	8.78
Total BOD_5, mg/l	143	10.9	12.9	112	8.3	12.7
Soluble BOD_5, mg/l	87	3.2	2.6	65	1.7	2.0
Wastewater Temperature ^{o}C (^{o}F)	11 (52)			10 (50)		
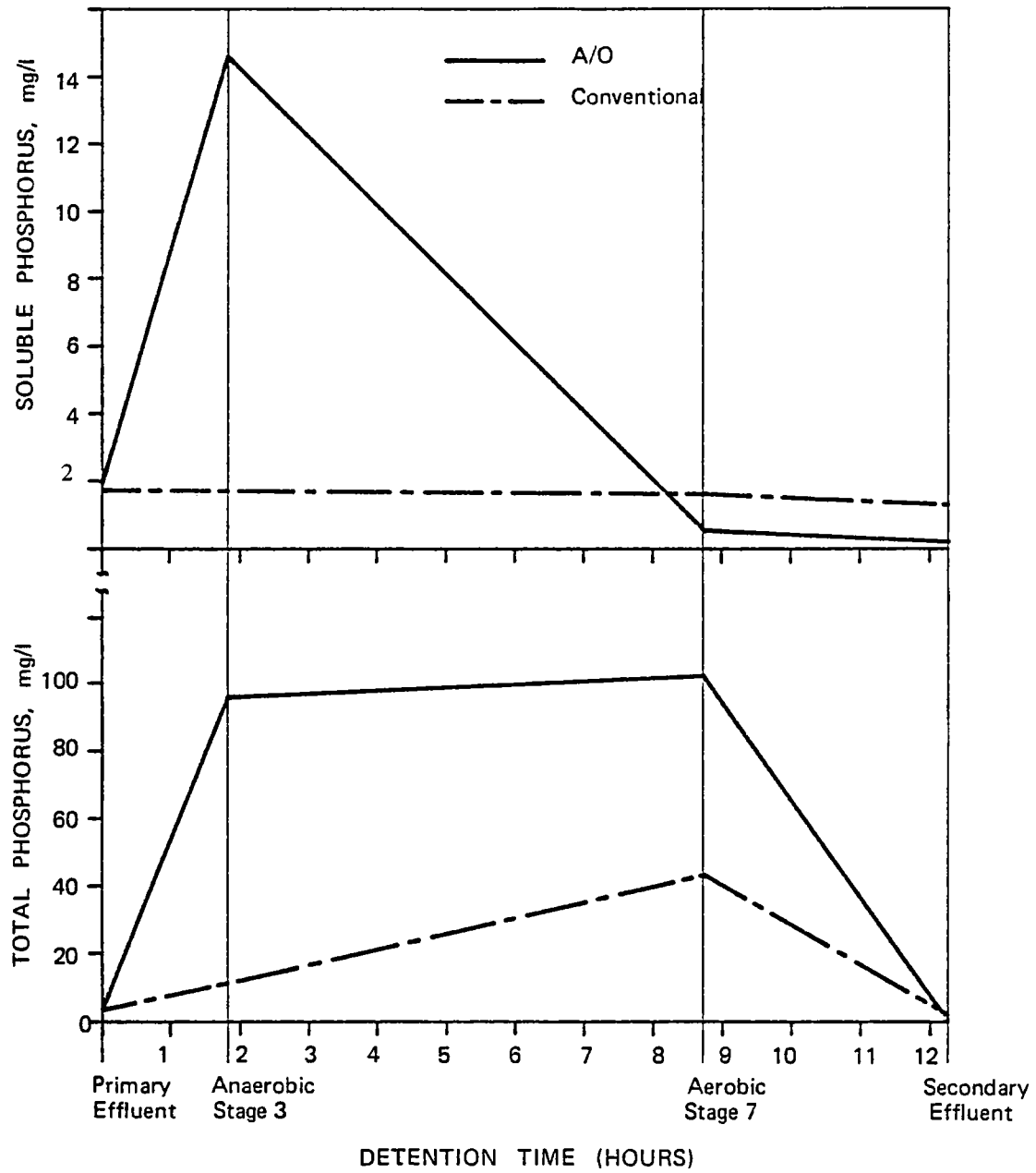
FIGURE 3

PHOSPHORUS PROFILE
NITRIFICATION WITH DIURNAL
FLOW VARIATION

The removal of BOD_5 was consistently good throughout the study period, with removal efficiencies normally above 90%. Nearly 70% reduction in soluble BOD_5 occurred in the anaerobic zone during some period.

SETTLEABILITY

The settleability of the activated sludge produced in the A/O and conventional trains was evaluated using the sludge density index (SDI).

SDI is defined as:

$$\frac{MLSS\ (\%)\ x\ 100}{\% \text{ Volume occupied by MLSS after 30 minutes settling}}$$

A well-settling sludge would have SDI between 1.0 to 2.5, while an SDI less than 0.5 indicates a poor quality sludge. The average SDI's for the A/O and conventional trains calculated for each of the four phases of study are presented in Table 3.

TABLE 3
SETTLEABILITY SUMMARY

PHASE	AVERAGE SDI	
	A/O	CONVENTIONAL
I	1.68	1.99
II	1.64	2.06
III	0.99	1.63
IV	1.07	1.63

The SDI's for both systems are within the range for a good quality sludge. The SDI's measured in the conventional trains, however, were consistently higher than those measured in the A/O trains.

NUTRIENT UPTAKE

Differences in nutrient uptake between the A/O and conventional systems were investigated. The ratio of BOD_5 removed to phosphorus removed in each system for the four phases of study was calculated and is presented in Table 4. A similar calculation could not be performed for nitrogen, since an unquantifiable amount may be lost due to denitrification.

TABLE 4
BOD_5 TO PHOSPHORUS RATIOS

PHASE	A/O SYSTEM	CONVENTIONAL SYSTEM
I	100 : 2.4	100 : 1.3
II	100 : 2.7	100 : 1.5
III	100 : 2.6	100 : 1.5
IV	100 : 2.3	100 : 1.4

BOD$_5$ to phosphorus ratios were very consistent within each system. A good deal more phosphorus is taken up to remove a given amount of BOD$_5$ in the A/O system than in the conventional system. It is interesting to note that more phosphorus was removed in the conventional trains than is frequently given as a biological requirement in the literature.

The total nitrogen and phosphorus contents of the biological solids of each system were compared as shown in Table 5.

TABLE 5
NITROGEN AND PHOSPHORUS IN BIOLOGICAL SOLIDS

PHASE	A/O SYSTEM		CONVENTIONAL SYSTEM	
	TKN/TSS, %	T-P/TSS, %	TKN/TSS, %	T-P/TSS, %
I	4.45	2.74	4.51	2.40
II	4.14	3.30	4.09	2.01
III	5.12	2.98	4.90	2.01
IV	3.63	2.64	3.46	1.76

Although there is little difference between the two systems in terms of TKN content of the activated sludge, the A/O sludge does contain a greater fraction of phosphorus. Through the review of related work by other researchers, it is observed that the concentration of phosphorus contained in the biomass is proportional to the concentration in the influent wastewater.

IMPACT OF RECYCLES

The nitrate content of the return activated sludge was not observed to adversely impact the A/O process. Average A/O return sludge nitrate concentrations for the four phases of study ranged between 5 and 12 mg/l as nitrogen depending on the extent of nitrification. However, mixed liquor samples at the end of the third anaerobic stage in each train normally contained less than 1 mg/l of nitrate nitrogen. Good phosphorus removal continued through periods of nitrification with high return activated sludge nitrage concentrations.

Process performance was also not observed to be adversely affected by the recycle of digester supernatant. Slight increases in primary effluent phosphorus and ammonia occurred as a result of this recycle during Phase II, although the average primary effluent phosphorus concentration observed during the Phase IV digester supernatant recycle period was lower than during the other phases. This was due to a dilution effect caused by snow melt and a great deal of precipitation during this period.

CONCLUSIONS AND RECOMMENDATIONS

Retrofit of one-half of Pontiac's East Boulevard Plant was accomplished in February 1984. The evaluation of the A/O process performance has subsequently been completed.

The retrofit operation was finished in two (2) months at a cost of approximately $50,000 for 13,250 cu meters per day (3.5 mgd) treatment capacity. With respect to the process variables which required investigation, the following conclusions are made.

Good phosphorus removal was attained throughout the study period, and remained good through periods of cold temperatures. Average phosphorus concentrations in the secondary effluent of the converted activated sludge trains were well below 1.0 mg/l for each of the four periods of study with various combinations of nitrification and anaerobic digester supernatant recycle.

The recycle of nitrate in the return activated sludge during periods of nitrification was not shown to adversely impact the phosphorus removal process. Good nitrification and phosphorus removal was attained concurrently.

Recycle of digester supernatant was also shown not to be detrimental to the A/O process. Slight increases in phosphorus and nitrogen loadings were observed as a result of the recycle, but good removals continued.

Taking into consideration the above observations, the successful operation of the A/O process in a cold climate has been demonstrated.

BACKGROUND TO THE SELECTION OF THE MODIFIED

BARDENPHO PROCESS FOR KELOWNA

G. M. STEVENS

CITY OF KELOWNA

BACKGROUND TO SELECTION OF THE MODIFIED BARDENPHO PROCESS
FOR KELOWNA

G.M. STEVENS
CITY OF KELOWNA

The residents of the Okanagan Valley have long been environmentally sensitive and one of the major attractions to the Valley has always been the lakes and their natural beauty. To harm the lake system has been the cry to arms for many and the political system has responded with commitments to preserving the lakes in their natural state.

In the late 1960's Algal Blooms on several areas of Okanagan Lake caused much concern and the joint Federal/Provincial Okanagan Water Basin Study was initiated to investigate many facets of Okanagan Valley life including environmental concerns. The study concluded the main body of water in Okanagan Lake was still oligotropic and phosphorus was identified as the limiting nutrient. Furthermore, several foreshore areas were identified as nitrogen limiting especially near the discharges of major creeks. The technology available at the time indicated control of phosphorus loadings would be the least expensive and most effective means of ensuring the oligotropic condition of Okanagan Lake. Reductions in both the point and diffuse sources of phosphorus were recommended and several programs to accomplish this were implemented.

The major point source of phosphorus was identified as the discharge from sewerage systems and the water basin study recommended 80% phosphorus reduction from these sources. The City of Penticton implemented phosphorus removal at their treatment plant and became one of the early leaders in advanced waste water treatment. However, the local concern for the condition of the lake pressured other communities for higher levels of treatment. The City of Vernon with some readily available farm land, adopted the concept of spray irrigation of its effluent and the cost for complete removal of effluent from the lake appeared at the time to be acceptable.

The choices for the City of Kelowna were far more difficult to make and the City embarked on an extensive search for treatment plant and land disposal combinations that would meet the no effluent discharge requirement. In 1975 a waste water management plan was commissioned to make recommendations on the treatment plant and land disposal options. Public involvement, technical committees, financial analysis all indicated the choice in Kelowna was not going to be as easy as the others. Consideration was given to the following treatment plant and land disposal schemes: (1) Primary, secondary and lagoon treatment for carbon removal, (2) Lime, alum, iron for phosphorus removal, (3) Ion exchange, Carbon denetrification and Ammonia Scrubbing for nitrogen removal, (4) Spray irrigation of vegetable and forage crops to forest rejuvenation, (5) Ground exfiltration at several locations. Cost projections for the complete removal of the effluent discharge ranged from $40 - $70 million which the City could not hope to afford.

The major obstacle to implementing the plan was the inability to produce an effluent of high enough quality at a reasonable cost. Most treatment plant options in the late 1970's were limited to secondary treatment plant designs with add-on physical and chemical processes which produced the tertiary

quality effluent. Plant costs for these options are such that no additional monies would be left for an effluent removal stage. It was at this time the modified Bardenpho process was introduced to the City as a treatment plant option that would produce a high quality effluent at a cost very similar to secondary treatment.

Kelowna as early as 1913, implemented the latest technology available to treat the City's sewerage. At that time Imhoff tanks were the answer to effective treatment and these were constructed on the existing treatment plant site. In the 1980's biological nitrogen and phosphorus is the latest in high-tech treatment plant options and the City again took a bold step and implemented this new approach. The objective of the plant remains to produce a very high quality effluent which would be capable of spray irrigation without buffer zone restrictions; however with the current provincial restraint program, this further stage of treatment has not been implemented.

DESIGN OF THE KELOWNA POLLUTION

CONTROL CENTRE

Patrick J. Leslie

Knight and Piesold Ltd.

DESIGN OF THE KELOWNA POLLUTION CONTROL CENTRE

By
Patrick J. Leslie
Knight and Piesold Ltd.

INTRODUCTION

The treatment plant at the Kelowna Pollution Control Centre is a five stage Bardenpho plant designed for the simultaneous biological removal of nitrogen and phosphorus in addition to carbonaceous material. The basic process train for the bio-reactor consists of an anaerobic fermentation zone, a primary anoxic zone, an aerobic zone, a secondary anoxic zone, and a re-aeration zone. The principles of the Bardenpho Process are covered in detail in other papers being presented at this seminar.

DESCRIPTION OF TREATMENT PLANT

The former treatment plant at the site was a conventional activated sludge plant with a design capacity of 130 L/s (3 U.S. mgd). Part of the flow passed through two trickling filters which were a legacy from a still earlier plant. These trickling filters and the old chlorine contact tank were demolished during the construction of the new Bardenpho plant. The inlet works, primary clarifiers, and secondary clarifiers were incorporated into the new plant. The aeration tanks from the former plant are being used for flow equalization of primary effluent.

A simplified flow diagram of the new Bardenpho plant is shown on Figure 1. The plant is designed for an average flow of 263 L/s (6 U.S. mgd), with a maximum hourly flow 2 times the average flow. Raw sewage enters the plant from a 915 mm (36 inch) diameter sewer and flows through conventional headworks consisting of a barminutor, grit chambers with grit collecting mechanism, and parshall flume for flow measurement. Low lift pumps raise the sewage to three primary clarifiers which are retained from the former plant, but are now operated in a high rate mode. The primary effluent flows to the Bardenpho biological reactor, which is the heart of the new system. The layout of the bio-reactor is shown diagramatically on Figure 2.

The design was based on the following sewage influent characteristics:

COD	-	375 mg/L
BOD_5	-	225 mg/L
Suspended Solids	-	200 mg/L
Total N	-	35 mg/L
Total P	-	7 mg/L

The following effluent standards had to be met:-

BOD_5	-	8 mg/L
Suspended Solids	-	7 mg/L
Total N	-	6 mg/L
Total P	-	2 mg/L

Flexibility of operation was a dominant design consideration, and is achieved by:

(1)   Dividing the biological reactor into two modules. This minimizes underloading of the plant during the early stages of operation, and will also allow one module to be shut down during periods of high temperature. (It also assists in plant optimization, as one module may run as a "baseline" module, while process variations are made to the second module.)

(2)   Each module is divided into twenty-one square "cells" which are separated in the flow path by "over and under" concrete baffles and weirs. The fermentation and anoxic

cells are provided with stirrers to keep solids in suspension, and the aerobic cells with turbine type aerators. The aerators are equiped with two-speed motors so that certain cells may be operated either as anoxic cells or aerobic cells for process optimization. The theoretical hydraulic retention time in each cell is about 1 hour.

Turbine type aerators were selected, with the air being supplied by centrifugal blowers, to provide the greatest degree of flexibility of operation and control. These allow optimum oxygen concentrations to be maintained without interferring with mixing characteristics. While the oxygen transfer efficiency of turbine aerators is lower than fine bubble diffuser systems, this is offset by their greater reliability and flexibility, and the lower manpower requirements for maintenance and operation. The total oxygen requirement for the plant is 340 kg per hour. The maximum power consumption at design flow will be 280 kW for the blowers, and about an additional 250 kW for the turbine aerators, stirrers, and recycle pumps.

Control of the air supply will be governed by DO concentrations in the aerobic zone. Six DO probes are located in each module. Readings from these probes are averaged by a micro-processor control system, which allows a bias factor to be applied to any of the probe readings by the plant operator.

Recycle from the aerobic zone to the first anoxic zone is effected by three low head axial flow pumps in each module, one pump in each module being stand by. Each pump has a capacity of 22 ML/d.

SRT control is achieved by wasting on a volumetric basis from the re-aeration zone.

Three new 26.0 m diameter circular secondary clarifiers supplement the original two smaller (15.2 m diameter) circular clarifiers, with the two smaller clarifiers having approximately the same capacity as one of the new larger clarifiers. The new clarifiers have a liquid depth at the walls of 4.5 m, and a sloping floor giving a liquid depth of 6.6 m at the centre. The bridges, which carry the scraper mechanisms, are three quarters of the clarifier diameter in length, and have rim drives. Two 60 L/s centrifugal pumps for each clarifier are used for the return sludge. The clarifiers were built to our own design, with mechanical equipment supplied by Passavant.

The clarified effluent flows to four dual media filters each having a filter area of 2560 m^2 with 200 mm depth of silica sand, and 200 mm depth of anthracite. The mechanism providing for continuous automatic backwash is carried on a travelling bridge spanning the filter beds. The mechanism is activated by a simple water level sensor as the filtering head, and consequently the water level above the filters, builds up to a pre-determined level, or by a timer.

After filtration the effluent is chlorinated, and then flows to the chlorine contact basin. This is designed to provide a chlorine contact time of 60 min., but provision was made in the design to increase this to 120 min. should the longer retention time be required by the regulatory authorities.

Primary sludge is pumped to a 556 m^3 gravity thickener. After thickening it passes through a disintegrator, and is then pumped to storage, where it is mixed with waste activated sludge in the sludge storage vault. An optional flow path for the thickened sludge is to screen it through a 2.5 mm screen, the screenings going to the sludge storage vault, and the fine fraction and liquid being pumped back to either the fermentation or first anoxic zone of the bio-reactor.

Waste activated sludge from the bio-reactor is pumped to two dissolved air flotation thickening units using variable speed positive displacement pumps. Air flotation thickening was selected because of the need to keep the sludge aerobic and prevent the release of phosphate. Magnetic flow meters measure the volume of sludge wasted, which provides the control for solids retention time. The sludge is thickened in these units to about 3 - 4% solids and then stored in the storage vault, which has

sufficient capacity to store approximately three days sludge production.

From the sludge storage vault, the sludge along with the thickened primary sludge is periodically trucked away to be mixed with waste hog fuel (wood chips) and then composted, using the windrow method. After composting for a period, the sludge is used as a soil conditioner and fertilizer by the City's Parks Branch.

PROCESS DESIGN

The process design for the plant was based on criteria developed by Barnard (1974) from experience with the design of similar plants in South Africa and on bench scale tests performed by Oldham (1979) at the University of British Columbia to study the performance of the process at cold temperatures. Sludge production at the various solids retention times was based on a model developed by Barnard et al (1972) and which has subsequently been modified by accumulated full scale plant information.

Previous experience has indicated that there are two important prerequisites for obtaining good phosphorus removal. The first is that the plant must operate in a plug flow mode. The division of the bio-reactor into cells separated by "over and under" baffles and weirs ensures good plug flow characteristics in the Kelowna plant.

Secondly the activated sludge must be subjected to a periods of anaerobiosis, and certain minimum levels of oxidation-reduction potential must be reached. In general, the greater the degree of anaerobiosis, the better will be the removal of phosphorus. To achieve good anaerobic conditions it is necessary to design the plant to severely limit the amounts of nitrates and dissolved oxygen in the anaerobic zone. Nitrates will be introduced into the anaerobic zone with the return activated sludge if denitrification is incomplete, and this will occur if there is insufficient carbonaceous matter in the influent wastewater relative to the total nitrogen concentration. Also excessive aeration will result in dissolved oxygen being recycled to the anoxic zones, which will interfere with denitrification and lead to insufficient removal of nitrates. Ideally, the DO concentration in the aerobic zone should be just sufficient to allow for complete nitrification, but this may be difficult to achieve, and in practice aiming for an ammonia concentration of just below 1 mg/L should suffice.

Dissolved oxygen may be introduced in the anaerobic zone if there are cascades or turbulent conditions in the influent channels or sludge return lines, and care must be taken in the design to avoid these. Over mixing by stirrers should also be avoided for similar reasons. At Kelowna a special gate was installed in the channel between the primary settling tanks and the bio-reactor to flood the outlet weirs of the primary tanks and minimize aeration. The return activated sludge pumps were connected directly to the underflow of the final clarifiers and piped directly to the bio-reactor to prevent any contact with air.

The Kelowna plant was the first full scale plant designed to operate in the colder regions of North America. The design therefore had to address two important differences in the characteristics of the raw wastewater compared with those typical of South African plants, namely lower temperatures and more dilute sewage.

It was estimated that the mixed liquor temperature in the Kelowna plant would have an average low temperature of $10°C$, but that this could fall to about $9°C$ for short periods. This was lower than the lowest mixed liquor temperature of about $11°C$ observed in plants in South Africa. Consequently, a pilot plant investigation funded by Knight and Piesold Ltd. and carried out under the direction of Dr. Oldham was undertaken at the University of British Columbia. This project involved operating a bench scale model of the process for over a year in a constant temperature room of the Environmental Engineering Laboratory, during which the mixed liquor temperature was progressively lowered from an initial $18°C$, at which operation was first stabilized, to $6°C$. The results of this investigation indicated that whereas nitrogen removal might

suffer at temperatures below $10^{\circ}C$, 90% removal of phosphorus was possible down to mixed liquor temperatures of $6^{\circ}C$. Even at the lowest temperature nitrification was not completely lost, and about 50% nitrogen removal was still achieved. The project also provided the necessary design information on reaction rate constants at the lower temperatures.

In the design of the plant, careful attention was paid to minimizing heat losses and maintaining the mixed liquor temperature. To reduce surface area, the liquid depth in the bio-reactor was set at 6 m. A turbine aeration system was selected that would eliminate cooling from spraying the liquid in the air, and the air pipeline from the blowers to the bio-reactor was insulated.

The weak sewages common in North America result in low organic carbon to nitrogen and low organic carbon to phosphorus ratios, both of which may be detrimental to good phosphorus removal. Provided the COD to TKN ratio is higher than 10:1 sufficient carbon will usually be available for removal of nitrates, but below this figure special design considerations may be necessary. Primary sedimentation will normally lower this ratio, as the carbonaceous matter in raw wastewater is predominately in the particulate form, whereas the nitrogenous material is mainly soluble. For this reason the primary settling tanks at Kelowna are operated in a high rate mode. The COD to TKN ratio of the primary effluent which is the influent to the bio-reactor, varies between 7:1 to 10:1. As it was anticipated that insufficient carbon would remain in the primary effluent to ensure good denitrification as well as phosphorus removal, provision was made to feed a portion of the supernatant from the primary sludge gravity thickener back to the bio-reactor. The sludge retention period in the thickener is such that acid fermentation takes place, but methane gas formation is inhibited. Subsequent operating experience with the facility has given convincing evidence that the short chain organics, (mainly acetates) which are produced during acid fermentation contribute significantly to the good phosphorus removal achieved by the plant. The importance of the forms of the carbon compounds entering the bio-reactor, and in particular the proportion of those which are readily bio-degradable, has been highlighted by recent research. The sludge thickener therefore also performs the function of a carbon modifier.

The design of the secondary clarifiers was conservative, as any solids washed out of the system may contain significant quantities of phosphate, which will then contribute to the phosphate content of the effluent. Sludge must be removed continuously from the clarifiers and the sludge blanket kept low enough to ensure that anaerobic conditions will not develop, with a consequent release of phosphates to the liquid phase.

GENERAL DESIGN CONSIDERATIONS

Construction of the facility necessitated addressing some interesting geotechnical problems, arising from the fact that the site is underlain by sedimentary and deltaic deposits. The floors of the new structures are located in a 3 m thick layer of soft silt, above and below which are layers of sand and gravel. This, together with the high water table, required the use of an extensive dewatering system combined with sheet piles during construction.

The total construction cost of the works was approximately $12,000,000.

REFERENCES

Barnard, J.L., Englande, A.J. and Eckenfelder, W.W. Design optimization for activated sludge and extended aeration plants. Advances in Water Pollution Research Sixth Int. Conf. Jerusalem, Permagon Press.

Barnard, J.L. (1974) Cut P and N without chemicals. Water and Wastes Eng., 11, 7, 33-36.

Oldham, W.R. and Dew, H.P. (1979) Cold temperature operation of the Bardenpho process. Presented at 14th Canadian Symposium on Water Pollution Research.

Barnard, J.L. (1982). The influence of nitrogen on phosphorus removal in activated sludge plants. Wat. Sci. Tech. 14, 1/2.31-45.

Barnard, J.L., Stevens, G.M., and Leslie, P.J. Design strategies for nutrient removal plant. Presented at IAWPRC Post Conference Seminar, Paris 1985.

Oldham, W.K. Full scale optimization of biological phosphorus removal at Kelowna, Canada. Presented at IAWPRC Post Conference Seminar, Paris, 1985.
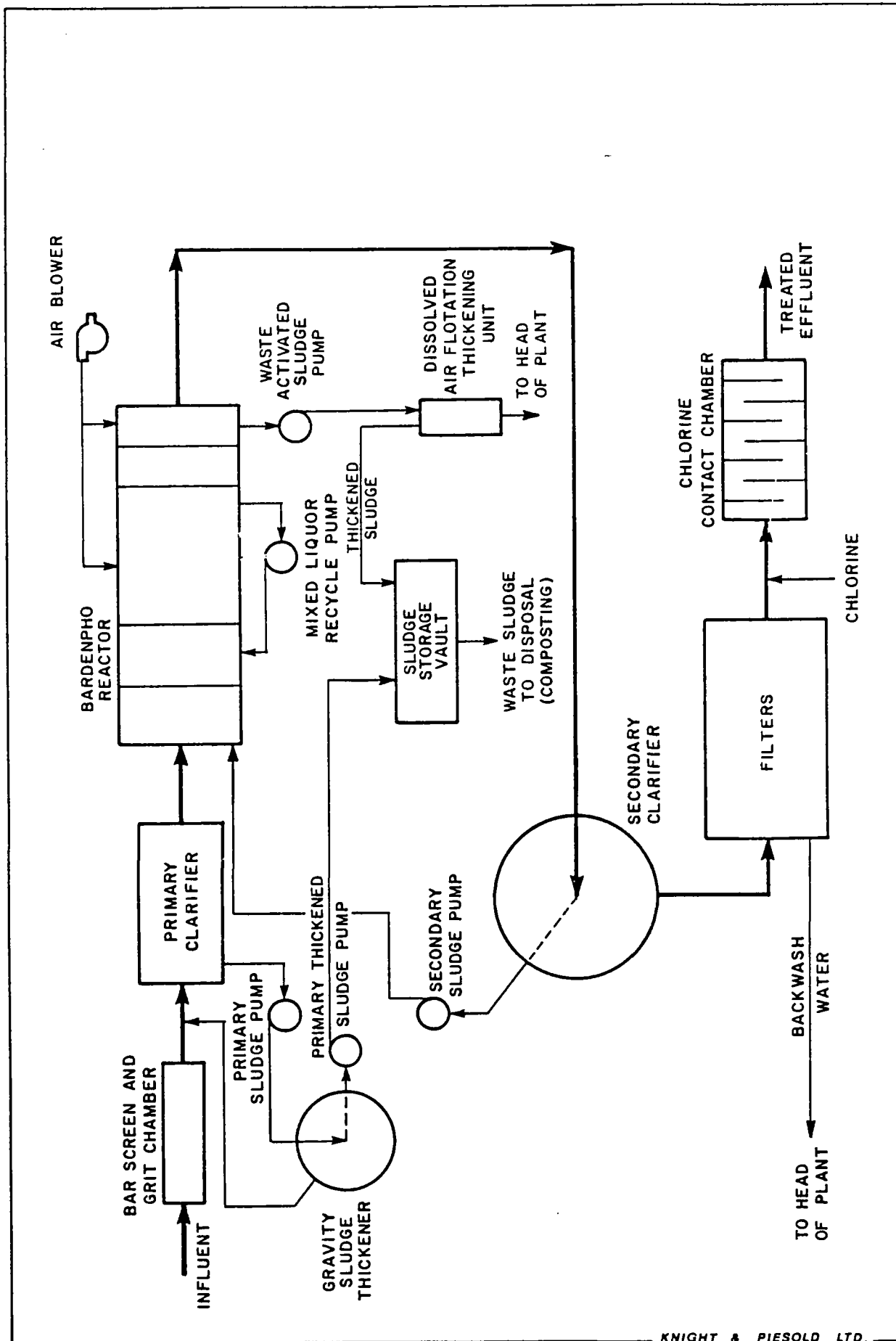
FIGURE I – SIMPLIFIED FLOW DIAGRAM OF
BARDENPHO PLANT
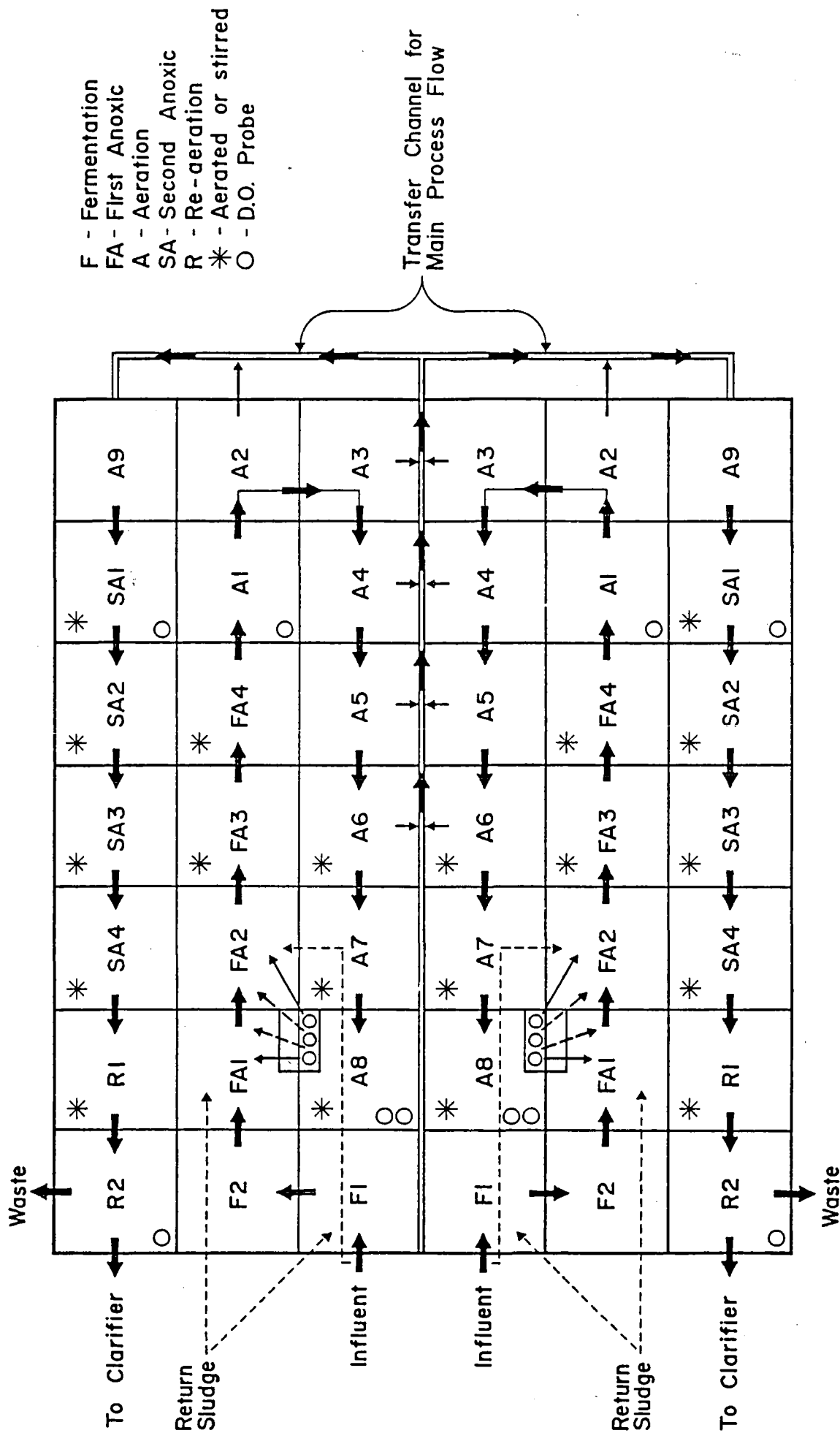
BARDENPHO REACTOR

FIGURE 2

OPERATING EXPERIENCES WITH THE KELOWNA FACILITY

G. M. Stevens

CITY OF KELOWNA

OPERATING EXPERIENCES WITH THE KELOWNA FACILITY

G.M. STEVENS
CITY OF KELOWNA

The operation of the first modified bardenpho facility in a northern climate has been challenging to say the least. However, the use of the biological process has been only one of those challenges. I would like to address some of the more important aspects of the equipment at the facility described in the previous paper and relate some of our operating experiences to date with this equipment. The design of this facility, while the process was still in the early years of its understanding, resulted in a complex layout with sometimes difficult operating procedures. However, it was important to provide flexibility and following the Kelowna experience future designs will be significantly simplified.

The use of microchip computer technology profoundly affected the design and operation of the Kelowna facility. The desire to operate the treatment plant 24 hours per day within tight operating limits and manpower constraints resulted in the complete automation of the treatment plant. Centralized control of the equipment and biological process has resulted in a concept whereby maintenance personnel ensure that the equipment functions and the computers ensure that the process functions. With daily checks of the process inputs by the staff, such as the dissolved oxygen probes, the computers assume their 24 hour per day duties of monitoring, controlling, data logging and alarming of all critical functions.

Logic control of the equipment motor control centres is performed by the Texas Instruments 5T1 System. Five logic control centres handle over 700 status inputs and outputs. Starting and stopping motors or other contact closure functions are programmed into the sequencers using the video programming unit which can either be located in the field or at a central work station. On-line status of all the programmable logic control (PLC) functions has simplified the trouble shooting of contact failures such as flygt bulbs, limit switches, or other open/close contacts.

Process analog, status, and alarm functions are all performed by Fischer and Porter DCI 4000 System. The flexibility of the system has time and time again proven its worth with changes occurring on site as the operational staff gain more knowledge of the key operating parameters controlling the process and equipment. Numerous operational functions change from summer to winter and the process computers have greatly simplified the monitoring of these changes and the manipulation of the process variables.

Comtinuous data logging, trending, and reporting is performed by the Westhead PDM 800 System. This 256K computer is directly connected with the process computers and can receive every 10 minutes up to 256 analog values for data logging and various report functions. Manual input of operator or laboratory values can be combined with the continuously logged data into calculations, trending, or hard copy reports. The ability to combine logged and manual data into various calculations has been invaluable in understanding the complex operations of the biological process.

The air delivery system is one example of the complete display and

manipulation of a complex control function made easy by the process computer. The process and PLC computers completely control blower starting or stopping functions and ultimately these decisions are based on the air requirements of the process. The Fischer and Porter equipment modulates the air supply to keep the dissolved oxygen levels in the aerobic zones at 2.0mg/L (+ 0.2mg/L).

The closed loop starts with the Leeds and Northrup dissolved oxygen probes. The design of the probe is based on a slightly different principle than other probes currently on the market. Within the membrane there is both an oxidation and reduction reaction and the probe measures this as it relates to the concentration of molecular oxygen. Thus the membrane does not have to support a continuous mass transfer of oxygen it only has to equalize the oxygen concentration on both sides of the membrane. The result is that a dirty membrane only reduces the response time of the probe and not the actual reading. The reduction of response time is not a factor in the control of dissolved oxygen in an activated sludge system as the daily fluctuations in flow and load occur very slowly. Cleaning of the probes twice weekly with a soap solution to remove the slime has proven satisfactory. Trending of the 24 hour data on the data computer combined with alarm limits for probe readings ensure accurate values.

Based on the dissolved oxygen readings the process computer averages the values and if desired can place a bias on a probe reading from a critical location in the process. This averaged dissolved oxygen level is then used in a cascading loop to determine the required air flow to each module. Orifice plates, pressure and temperature sensors convert the air flows to standard air flow measurements and the air flow is increased or decreased by pneumatic valves.

The air demand to each module is supplied from the common blower header. The four centrifugal blowers are modulated to produce a constant pressure over their 40% turn down capabilities and the process computer totals the desired air requirements of both modules to determine the number of blowers required.

The process computer also controls the movement of sludges on site from both the gravity sludge thickener and the wasting of activated sludge from the bioreactor. The operators daily measure the volume of sludge in the gravity thickener and enter the desired volume for transfer into the process computer. Through the use of magnetic flow meters the process computer then starts and stops the desired pumps based on the total sludge flow. The same principle is used for wasting activated sludge. The process computer will move a pre-determined quantity of sludge from the two modules to the dissolved air flotation thickeners and when the required total for each module is reached the pumps will shut down until the next day. The control of sludges on site has allowed consistent and controllable sludge retention times for both locations.

The expansion of the facility included an addition to the raw sewage pumping station. The previous arrangement used a magnetic clutch to vary the output of a pump and modulate it according to the wet well level. The new system uses a standard motor with a variable frequency drive which is modulated according to the wet well level. This results in the raw sewage pump station power requirements matching the raw sewage flow without the continuous starting and stopping of pumps.

With the introduction of variable frequency drive to the plant the concept has expanded to solve a problem with the bioreactor stirrers in the anoxic zones. Despite the low power inputs to the modules the stirrers are

producing some vortexing and aeration of the de-nitrification zones. The variable frequency drive equipment will reduce the vortexing and will cost approximately $30,000 to install and result in annual savings of $12,000 per year in power costs. This will allow us to fine tune the power inputs to the anoxic zones such that the stirrers will suspend the organisms and minimize unwanted air entrainment.

The demands for nutrient analysis have resulted in the purchase of auto-analyzer equipment for the laboratory. Two simultaneous analysis are made from one filtered sample every 1.5 minutes. This has allowed the monitoring of a large number of stations several times per week and subsequently very close supervision of the process for the evaluation procedure currently underway.

OPERATING EXPERIENCES WITH THE KELOWNA POLLUTION

CONTROL CENTRE--TREATMENT RESULTS

W. K. Oldham

Department of Civil Engineering

University of British Columbia

OPERATING EXPERIENCE WITH KELOWNA POLLUTION CONTROL CENTRE — TREATMENT RESULTS

W.K. Oldham
Department of Civil Engineering
University of British Columbia

INTRODUCTION

The Kelowna Pollution Control Centre (KPCC) has been operating as a Modified Bardenpho (Phoredox) plant since May 1982. It was constructed as two completely separate, parallel modules, so that the process downstream of the primary clarifiers can be operated in two modes simultaneously. It is thus possible to test two different physical configurations or operational procedures without the added effects of variability of sewage characteristics with respect to time being superimposed upon the test results.

For the first six months of operation, the entire sewage flow was routed through one module of the new facility. Under this operational mode, the hydraulic loading on the bioreactor and secondary clarifiers was somewhat in excess of the design load of 11.5 ML/d per module. The treatment results for that period of operation have been previously published (Oldham and Stevens, 1984).

Since January 1983, the facility has been generally operated in the parallel reactor mode, with intermittent periods of intense testing when significant operating differences between the two modules were being evaluated. For the period May 22, 1984 to November 9, 1984 the plant was converted back to single module operation to allow tests to be made on oxygen transfer rates in tap water placed in the non-functioning module of the plant.

This paper describes the general operating results during the 1983-84 period, and comments on the effects that various influent characteristics and operating parameters have an effluent quality.

PLANT EXPECTATIONS

On the basis of early South African experience (prior to 1980) with full-scale Modified Bardenpho plants and cold temperature laboratory studies at U.B.C. (Oldham and Dew, 1979), negotiations between the City of Kelowna and B.C. Ministry of Environment produced effluent quality requirements as shown in Table 1.

TABLE 1   PROVINCIAL PERMIT REQUIREMENTS FOR KPCC EFFLUENT

Parameter	Allowed Value mg/L
BOD_5	8
SS	7
Total N	6
Total P	2

It was considered that the quality outlined in Table 1, when combined with effluent disinfection, would ensure the following:

1) at least short-term protection for Okanagan Lake in terms of minimizing eutrophication; and

2) the ability to consider a wide range of crop irrigation schemes for which the effluent could be safely used.

Process design for the facility was carried out on the basis of raw sewage characteristics presented previously in this session.

GENERAL EFFLUENT QUALITY

During the 1983 and 1984 calendar years, the KPCC was being used to experiment with respect to the effects that a number of operating variables would have on the bioreator effluent quality. However, there was still the general over-riding requirement that the total plant effluent (the combined effluent from the two modules after filtration) should meet the Provincial effluent quality stipulated in Table 1.

Over that two-year period of time, the measured characteristics of the bioreactor influent were as shown in Table 2. Based on these figures, the process was functioning at about 54 percent of its design hydraulic flow during the two-module operating period. The plant loadings with respect to the other influent characteristics shown in Table 2 were something less than 50 percent of design during the two-module operating period. However, exact comparisons cannot be made because of the lack of analytical data for raw sewage inflow to the KPCC.

TABLE 2   Bioreactor Influent Characteristics.
(January 1983 - December 1984)

Characteristic	Median Value	Lower 5%	Upper 5%
Daily Volume	12.4 ML	10.4 ML	16 ML
COD	195 mg/L	150 mg/L	275 mg/L
Total-P	4.5 mg/L	3.3 mg/L	5.8 mg/L
Ortho-P (sol.)	3.8 mg/L	3.0 mg/L	4.3 mg/L
TKN (Total Kjeldahl Nitrogen)	24.5 mg/L	19.0 mg/L	33.5 mg/L
NH_3-N	17.5 mg/L	15.0 mg/L	21.1 mg/L
NO_3-N (sol.)	0.4 mg/L	0.1 mg/L	1.4 mg/L

During the single-module operating period (day 507 to day 678 on Fig. 1), the average inflow rate was 14.0 ML/d, or 120 percent of the design flow of 11.5 ML/d per module. Under this flow loading, the influent loadings of the quality parameters shown in Table 2 will be essentially at process design values, given the expected efficiency of a primary clarifier with respect to these parameters.

During the operating period reported on herein the general operating conditions within the plant were as follows:

a) mixed liquor suspended solids (MLSS)
   - 1100 to 2200 mg/L during two-module operation
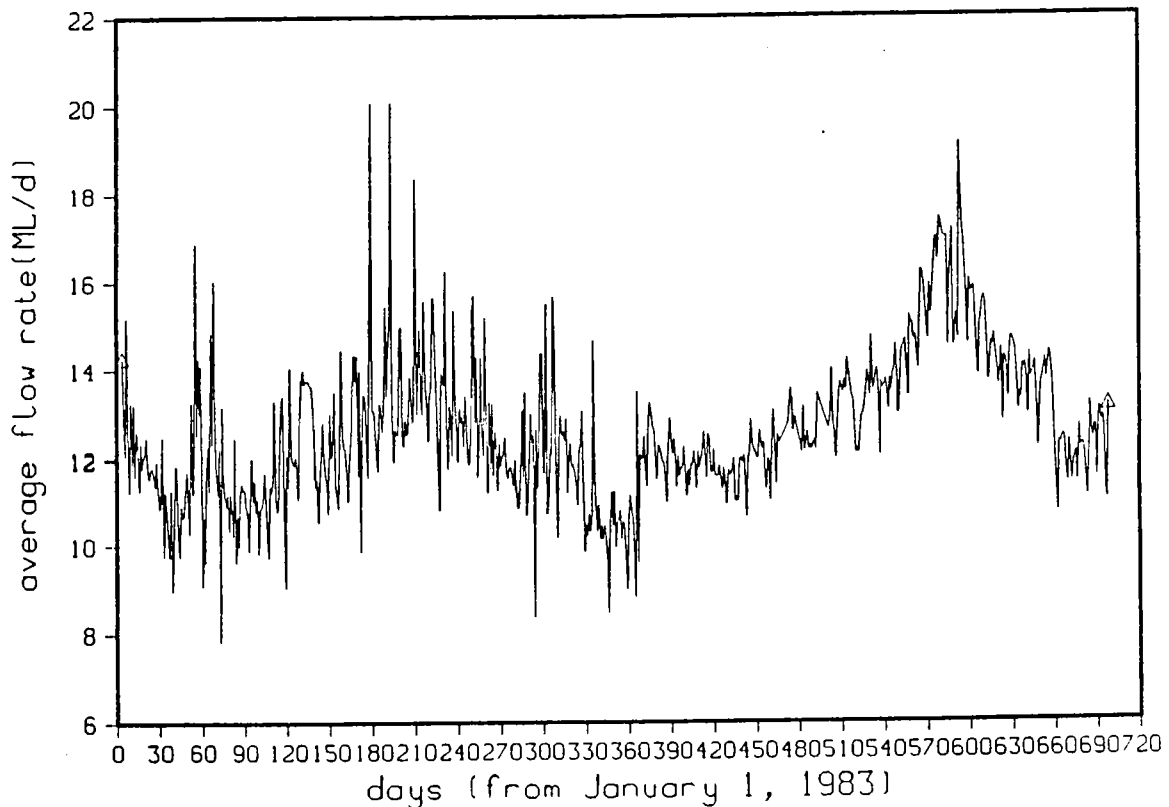   - 1700 to 2750 mg/L during single-module operation.

Fig. 1 - Influent flow rate.

b) total solids retention time (SRT)
    - 21 to 35 days.

c) aerobic mass fraction (fraction of biomass under aerobic conditions)
    - 0.45 to 0.6.

d) one-half of the secondary clarification facilities was in use (hence clarifiers were operating at, or above, design capacity).

e) return sludge was pumped at a constant rate of 110 to 120 L/S, which translates into an average return sludge rate of just under one times influent flow.

BOD and Suspended Solids Removal

The removal of the conventional impurities, BOD and SS, cannot be numerically quantified. Because the KPCC operates as an extended aeration plant (long HRT and SRT), the soluble biodegradable organics remaining in the secondary clarifier effluent is very low, and in fact, not measurable by conventional BOD testing procedures. By passing such a clarified effluent through a dual-media filter (see Fig. 2), a final effluent is obtained which has both a total BOD and a total SS less than the normal detection limits. The testing that has been done shows that BOD is continuously less than 10 mg/L and SS is less than 4 mg/L.

Nitrogen and Phosphorus Removal

Cumulative frequency plots of all the final effluent nutrient data acquired during 1983 and 1984 were prepared, and the information is summarized in Table 3. There are some apparent discrepancies between the ortho-P
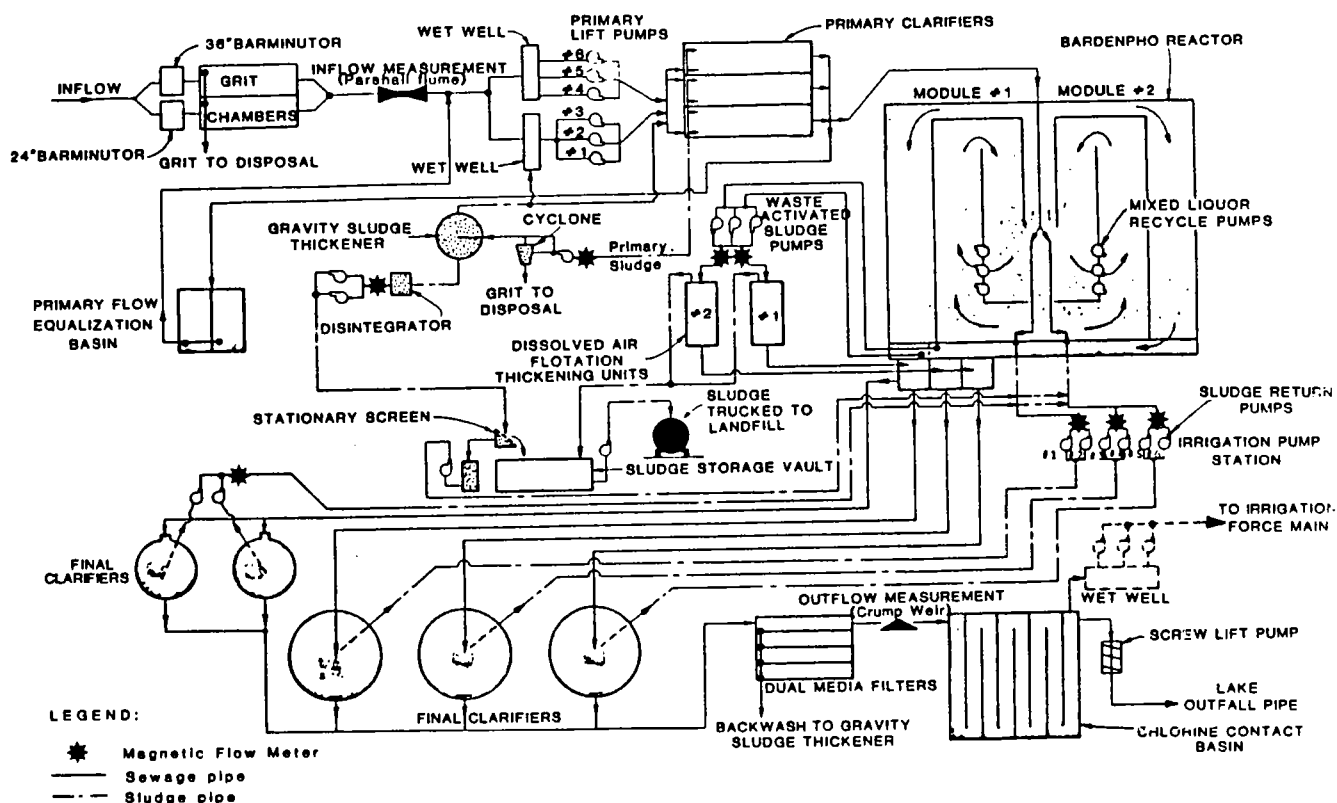
Fig. 2 — Plant schematic.

and total-P data, and between the ammonia-N and TKN data, especially at the upper 5 percent values. The reason for these discrepancies lies in the number of analyses performed — both total-P and TKN analyses were performed infrequently, while the ammonia-N and ortho-P analyses were performed at least twice per week, with higher frequencies during periods of poor perform-ance.

Since Table 3 shows that total-P and ortho-P values for the same cumula-tive frequency are very similar, and since the upper cumulative frequency concentrations are actually higher for ortho-P (as explained above), it is considered safe to say that Fig. 3 provides a worthwhile estimate of the percentage compliance with the P limit for the KPCC discharge. That is, an effluent phosphorus concentration of 2 mg/L or less has been met 90% of the time during 1983/84. When the total nitrogen situation is observed, the ordinate values on the cumulative frequency plots for TKN and nitrate-N

TABLE 3   FINAL EFFLUENT NUTRIENT CONCENTRATIONS
(January 1983 — December 1984)

Characteristic	Median Value		Lower 5%		Upper 5%	
Total-P	0.8	mg/L	0.2	mg/L	1.8	mg/L
Ortho-P	0.77	mg/L	0.15	mg/L	2.25	mg/L
TKN	1.5	mg/L	0.2	mg/L	4.0	mg/L
NH_3-N	0.1	mg/L	<0.1	mg/L	6.0	mg/L
NO_3-N	1.8	mg/L	0.8	mg/L	4.2	mg/L

cannot simply be added, because periods of high nitrate-N in the effluent are generally associated with low TKN values, and vice-versa. At present, all that can be said is that the median total-N concentration is about 3 mg/L (TKN plus nitrate-N), while 6 mg/L total-N is exceeded much less than 10 percent of the time (if high nitrate-N and high TKN occurred simultaneously, 10 percent non-conformity with the effluent N requirement would be correct).
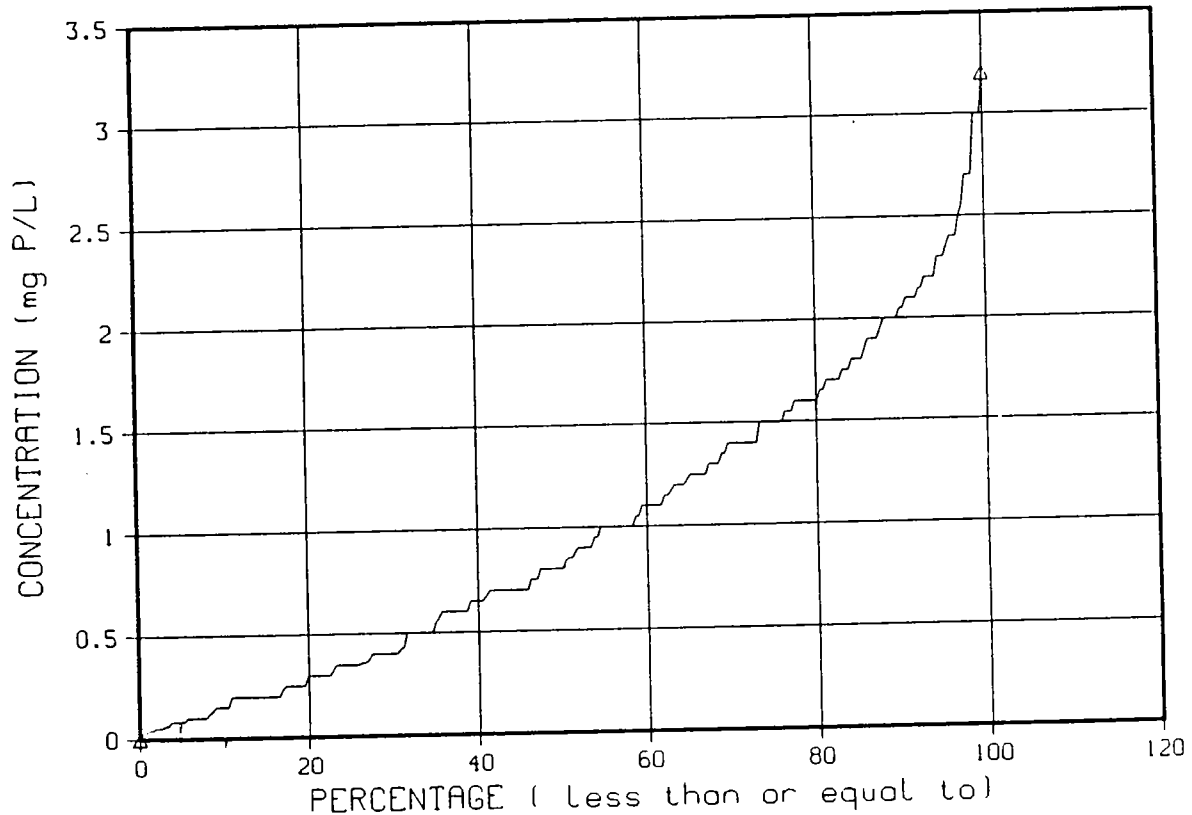

Fig. 3 - Cumulative frequency plot, 1983/84, effluent ortho-P.

To set the process success rate in better perspective, it is worth inspecting the effluent characteristics for the 170 day period in 1984 when the total inflow was being treated in one module (i.e. the process was operating at full design load). Table 4 provides that information.

TABLE 4   FINAL EFFLUENT CONDITIONS DURING SINGLE MODULE OPERATION
(May 22 - November 9, 1984)

Characteristic	Median Value	Lower 5%	Upper 5%
Daily Volume	14.0 ML	12.0 ML	17.0 mg/L
Ortho-P	1.1 mg/L	0.08 mg/L	1.75 mg/L
NH_3-N	0.04 mg/L	0.03 mg/L	0.75 mg/L
NO_3-N	2.00 mg/L	1.20 mg/L	3.4 mg/L

In comparing the data in Tables 2,3, and 4, the following points are evident:

1) The daily flow rate was 15 percent higher during the period of single-module operation than during the whole two-year test period. Additionally, the median total-P concentration in the bioreactor influent during single-module operation was 4.8 mg/L, as compared to the 4.5 mg/L shown in Table 2.

2) Median effluent ortho-P during single-module operation was not quite as good during the single-module operation (1.1 mg/L vs. 0.77 mg/L); however, the severity of excursions on the high side were not nearly as great (see Upper 5% values in Tables 3 and 4). A more consistent effluent quality would be expected during single-module operation, because no purposefully detrimental process changes were implemented during this period of time.

3) The summary of NH_3-N data in Table 4 indicates that better nitrification was achieved under single-module operation. The reason for this difference is made obvious by looking at Fig. 4. The loss of nitrification during February and March of 1983 skewed the data used for constructing the NH_3-N entries in Table 3.

4) Nitrate-nitrogen data in Figs. 3 and 4 give similar conclusions to that for phosphorus — the median value isn't quite as good under single-module operation, but the severity of upward excursions is less.
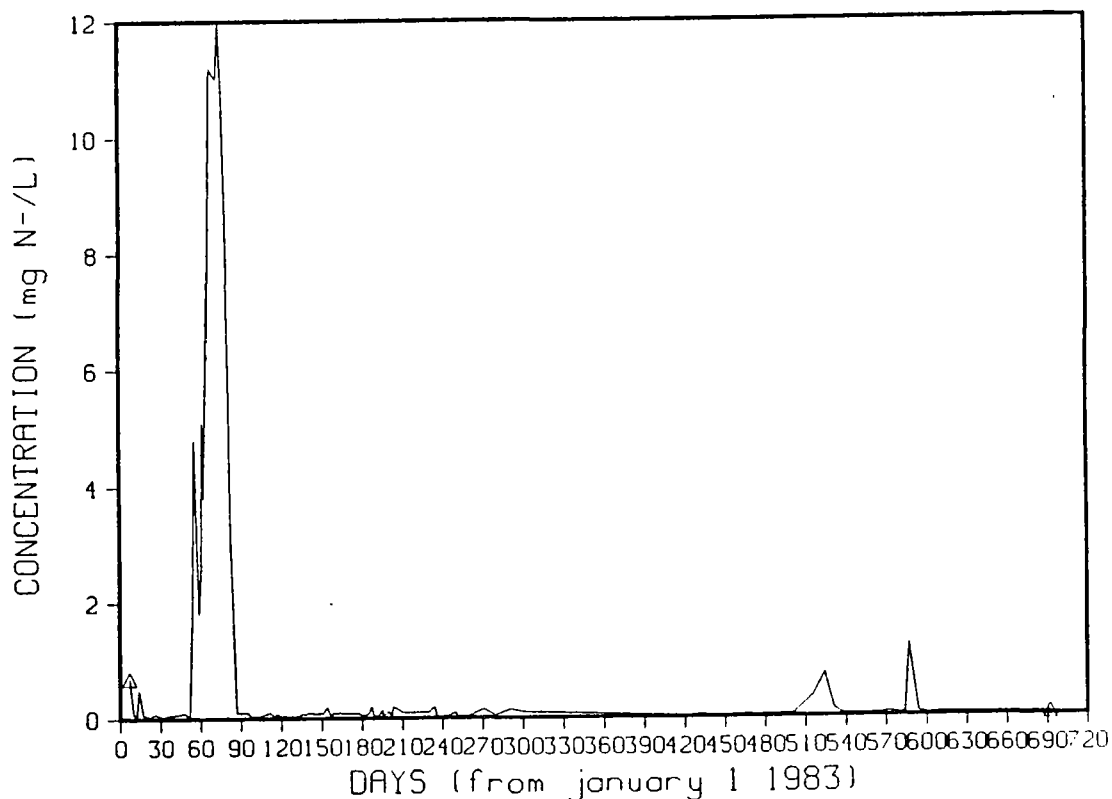
Fig. 4 - Effluent ammonia-N.

The overall comparison of single-module vs. dual-module operation of KPCC over the past two years indicates that average removals are essentially the same, but that percentage compliance with the Provincial regulations has actually been better under full-load conditions. As stated earlier, such is expected because there were times during the dual-module operating period when the operating conditions in one module were purposefully altered with the intention of making the process less efficient.

FACTORS AFFECTING PHOSPHORUS REMOVAL

Although the KPCC is required to produce an effluent with low concentrations of various constituents (Table 1), phosphorus removal only is being considered at this technology transfer session. Hence, the remainder of this paper will deal exclusively with that process technology, and will bring in data on other aspects of the process only when it directly influences P removal.

The biological phosphorus removal process has been the subject of much investigation over the past ten years. Although the phenomenon has been observed sporadically in activated sludge plants since the 1960's, specific research on understanding it has really only occurred since about 1974. Laboratory-scale, pilot-scale, and full-scale processes have been operated in South Africa, Europe, and North America with varying degrees of success. A set of operating conditions which provide very good results in one location for a certain period of time do not necessarily give nearly the same success with another sewage, or at another time with the same sewage.

It is generally recognized that the success of extracting phosphorus from solution in raw sewage, and holding it as a part of the waste activated sludge solids is highly dependent upon developing a sufficient concentration of specific bacteria in the activated sludge biomass. The trick, then, is to provide environmental conditions that favour the growth of such organisms over the growth of the normal heterotrophic population in an activated sludge basin. In short, we need a relatively large number of process-specific organisms that can uptake and store phosphorus in large excess over that which is required for normal metabolic activities.

There is still a lot of controversy over how many different species of organisms there are that can be cultivated to take on this function in a modified activated sludge plant, and by what specific biochemical pathways the transformation occurs. However, such lack of knowledge has not prevented researchers from finding out a lot about environmental conditions required to facilitate the growth and efficient functioning of such organisms.

Generally, it is postulated that the following conditions are necessary to allow significant biological removal of phosphorus to occur.

1) The presence of a simple organic substrate in the influent sewage is necessary. It is postulated that it can be quickly transported across the cell membrane of the bio-P organisms, and subsequently stored in a high energy state as polyhydroxybutyrate (PHB). During this carbon storage process, ortho-phosphate is released from the cell (Comeau, 1984).

2) The presence of a truly anaerobic zone is necessary in the process train, so that the desired bio-P organisms can make use of the simple substrate without undue competition from normal facultative organisms.

3) A long enough SRT is required to ensure that the desired organism can increase in number, and not be washed out in the waste activated sludge (WAS) as fast as they are formed.

4) As is the case with any living organism, it is necessary to have a low enough concentration of potentially toxic compounds in the bioreactor so the desired biochemical reactions are not affected.

If the above conditions can be maintained, then the bio-P organisms can maintain a preferential metabolic position in the activated sludge biomass. When the biomass is subsequently moved from the anaerobic zone to the aerobic zone, it is capable of very fast uptake of phosphorus as an alternative and more advantageous energy storage mechanism. It is the stored PHB that allows this P storage mechanism to be activated.

In translating these perceived essential conditions into plant design and operating decisions, the following list arises.

1) Plant inflow should not be fresh, as the presence of DO is detrimental to the desired function of the anaerobic zone.

2) If nitrification occurs in the aerobic zone of the bioreactor, the denitrification of at least the return sludge should be practised, so that the mass of nitrates returned to the anaerobic zone is minimized.

3) The presence of short-chain fatty acids (acetate and propionate) entering the anaerobic zone are beneficial, if not necessary, to the process.

4) Waste activated sludge must be handled in such a way that no resolubilization of P occurs during thickening in situations where the liquid stream is returned to the main treatment process.

During the two years of plant operation for which detailed analytical data has been obtained, it has been possible to look at the effects that some of the above conditions have had on effluent quality. Some of the comparisons have been carried out by operating the two parallel process modules under distinctly different conditions, while others have been made by looking at temporal changes in the operation of the process.

Selection of Test Conditions During Dual-Module Operation

The main criterion considered in selecting the operational variables to be tested was the anticipated impact on phosphorus removal, even though the Phoredox process is intended to give high removals of both nitrogen and phosphorus. To achieve good phosphorus removal, it has been held by many researchers that the return sludge must pass through an anaerobic zone (Davelaar et al., 1978; Nicholls and Osborn, 1979; Barnard, 1982; Siebritz et al., 1983). Additionally, for that anaerobic zone to be highly effective in conditioning the return sludge for substantial phosphorus uptake, it appears imperative that the flow entering that zone has a combination of a high concentration of simple organics and/or a low concentration of nitrates (Fuhs and Chen, 1975; Ekama et al., 1983; Comeau, 1984).

Since the degree of anaerobiosis attained in a given anaerobic zone is inversely proportional to the load of oxygen and nitrate entering that zone, and since the main source of these components in the Kelowna plant is the return sludge flow, it was identified as one of the process variables to investigate. The return sludge pumping rate for each module has been set at 60 L/s, because operating experience has shown it to be successful in preventing the formation of a significant sludge blanket in the clarifier (thus ensuring a minimum of P release in the return sludge). One way of reducing the mass loading of nitrates and oxygen to the anaerobic zone without reducing the total return sludge flow rate is to split that flow into two streams — one portion to enter the anaerobic zone, the other to discharge to

the first anoxic zone (see Fig. 5). It has been postulated by Barnard (1983) that the exposure of only a portion of the biomass to improved anaerobic conditions on each pass through the system will increase overall P removal. Such process operation was therefore selected for study at Kelowna, with one module operating on a 50/50 split of return sludge between anaerobic and anoxic zones, while the other returns 100% of the sludge to the anaerobic zone.
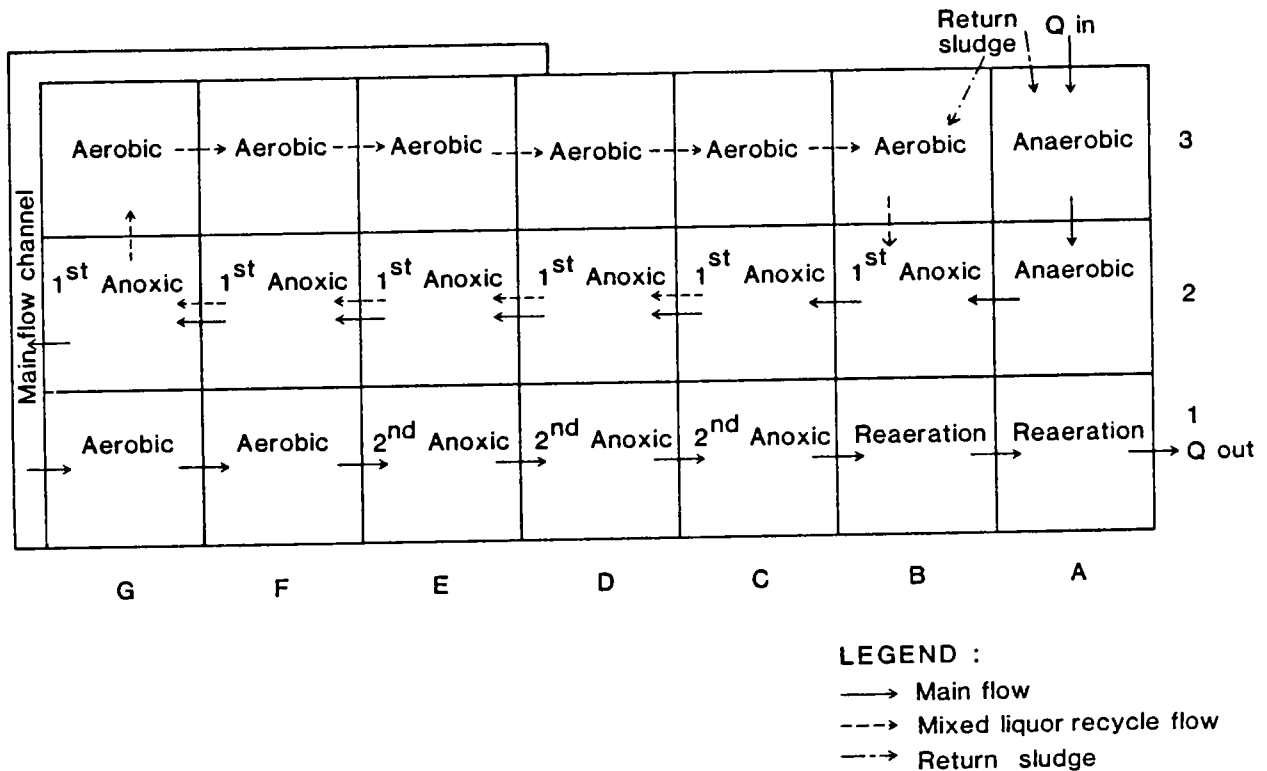

Fig. 5 - Bioreactor module configuration.

The importance of easily biodegradable organics in the anaerobic zone has been recognized by many researchers. It has been postulated (Ekama et al., 1984) that these organics can easily be converted to acetates by facultative organisms in the anaerobic zone, and thus stimulate the P-release and polyhydroxybutyrate (PHB) storage phenomena described by Fuhs and Chen (1975). One method of accelerating the process of acetate production is to anaerobically store the sludge from primary clarifiers for sufficient time to allow the growth of acid formers to occur, while still restricting the growth of methane formers. Since the primary sludge thickener at Kelowna provided an acceptable location for acetate production to occur, the impact of the thickener supernatant on the P removal efficiency was chosen as the second optimization investigation (see Fig. 2).

Normally, thickener supernatant is returned through the primary clarifiers where, along with the clarified sewage inflow, it is split evenly between the two modules. During the thickener supernatant testing program, it was pumped directly to the anaerobic zone of the module selected to receive the benefit of the added easily biodegradable organics.

PLANT PERFORMANCE UNDER VARIOUS CONDIITONS

Within this section there are discussions of the results of the specific dual-module testing configurations, plus some comments on other plant conditions which apparently had some effect on final effluent quality.

Point of Entry of Return Sludge

Two separate tests of the effect of splitting return sludge (RS) between the anaerobic (fermentation) zone and the first anoxic zone have been carried out to date. The general operating conditions within the plant during these periods are summarized in Table 5. The only operating parameter that was substantially different for the two tests was mixed liquor temperature, with both measured ranges being indicative of winter, rather than summer, conditions.

TABLE 5 PLANT OPERATING CONDITIONS DURING SPLIT RETURN SLUDGE TESTS

	Test 1	Test 2
Dates	Jan.-Feb. 1983	April 1983
SRT(d)	24	25
MLSS (mg/L)(avg.)	1450	1600
Mixed Liquor T (°C)	10-11	13-14.5
Inf. TKN/COD (avg.)	0.12	0.12
Inf. Q (L/s)(avg.)	120	127
Return Sludge Flow (L/s)	120	120

With respect to overall performance, the data plots for "end reaeration zone" (cell 1A on Fig. 5) on Figs. 6, 7, 8, and 9 indicate strongly that there are no significant differences in either dissolved phosphorus or nitrate in the bioreactor effluent, regardless of the return sludge operating mode. Although no specific plots for NH_3-N are available for "end reaeration zone", reference to Fig. 4 will show that the final effluent NH_3-N is consistently very low during the test period. Hence the two bioreactor effluent streams must also be uniformly low, as there is no opportunity for further nitrification to occur between the end of the reaeration zone and the sampling point for the final effluent.

Within the bioreactor, some apparent differences in module performance are notable. For example, there is greater phosphorus release in the fermentation (anaerobic) zone of the North module than in the South module (see Figs. 7 and 9). However, this appears to be a function of physical or operational differences other than the split of return sludge, as the higher concentration is consistent regardless of the return sludge entry point. When considering the actual release of phosphorus that occurs per unit volume of return sludge entering the fermentation zone of the bioreactor, the effect of splitting the return sludge flow is evident. A phosphorus mass balance calculation on the streams entering and leaving the fermentation zone shows that a higher unit release of phosphorus is achieved under the split flow regime (see Table 6), but much more data are required to establish a relationship.
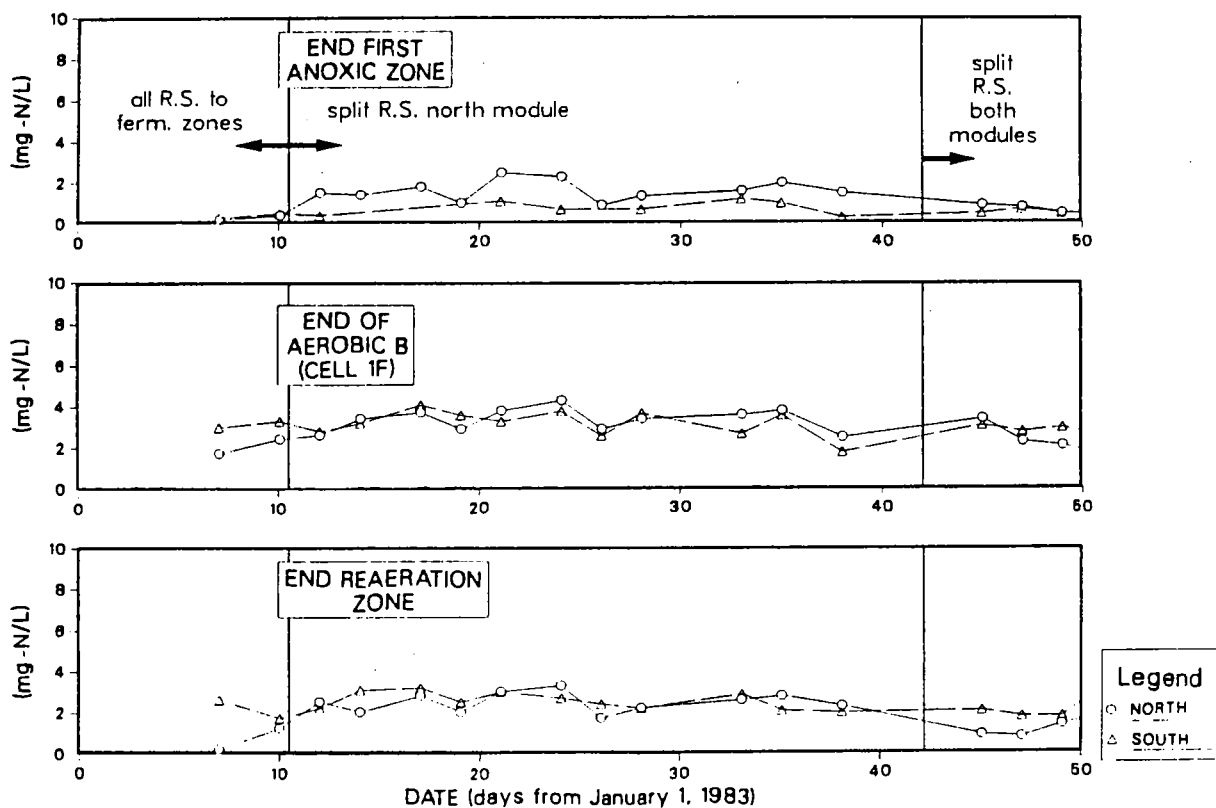
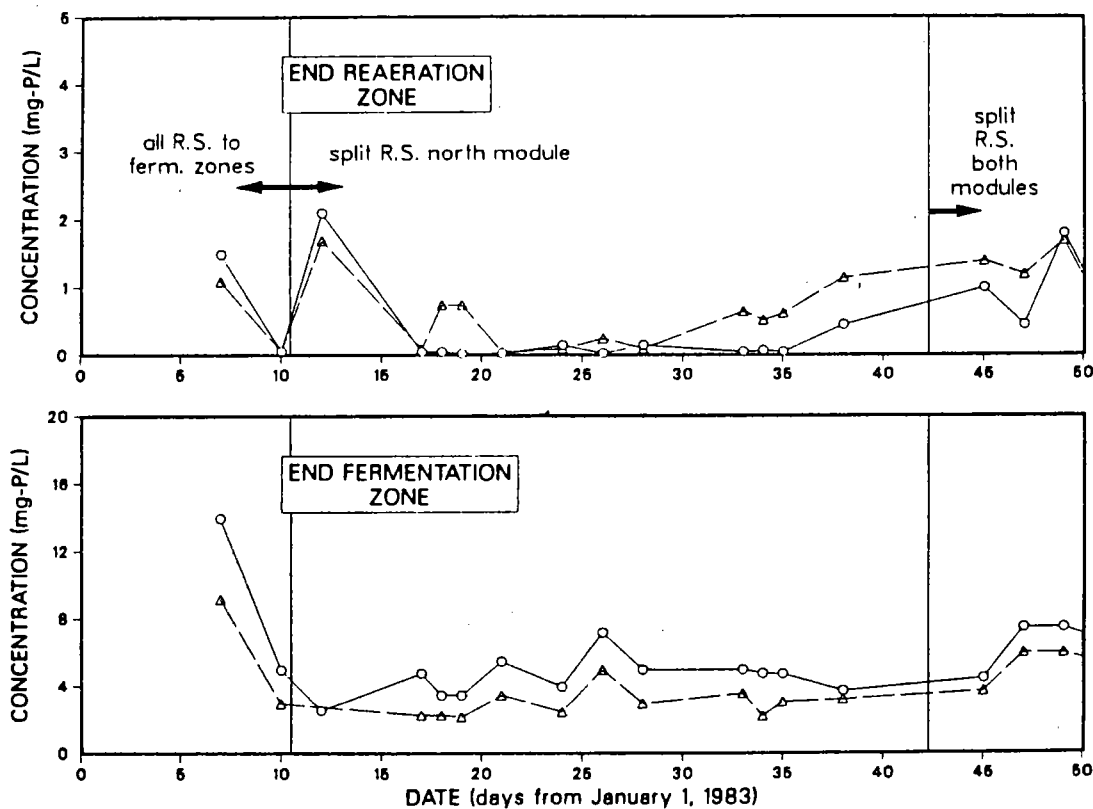Fig. 6 - Return sludge split test #1:nitrate-nitrogen.

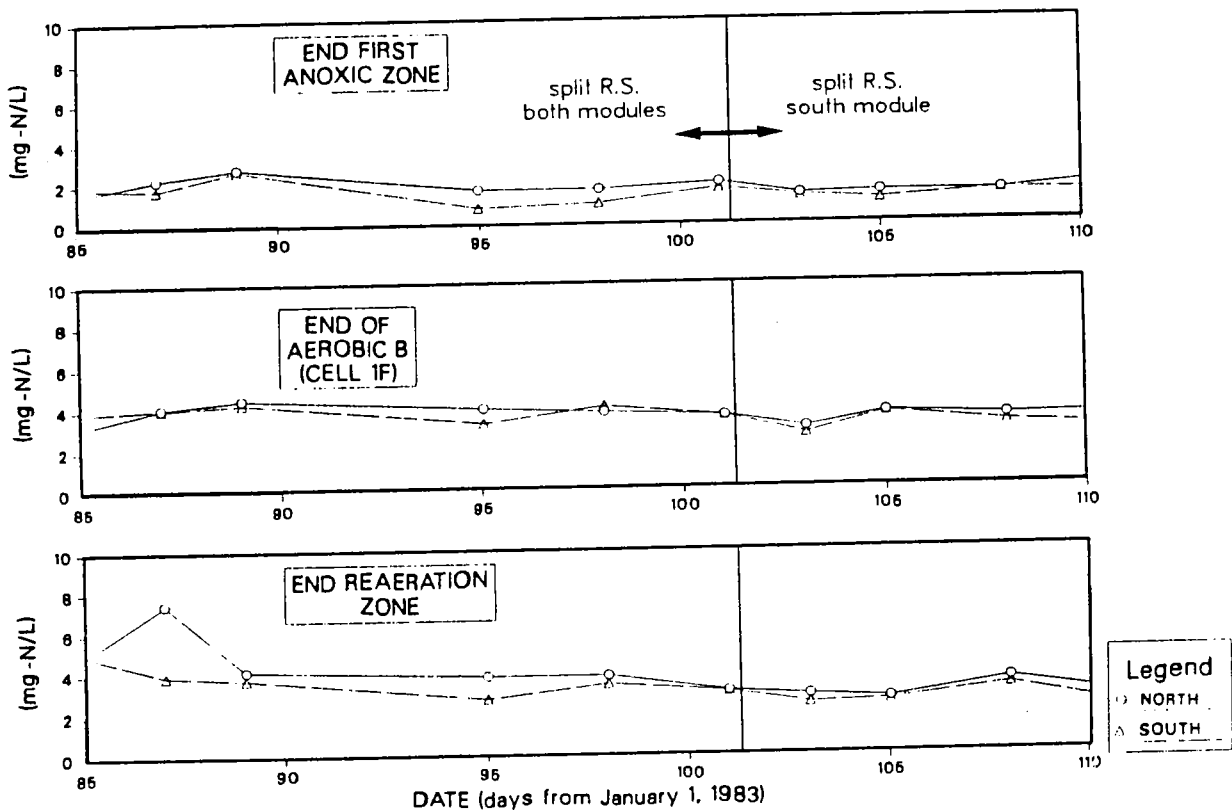Fig. 7 - Return sludge split test #1:orthophosphate.

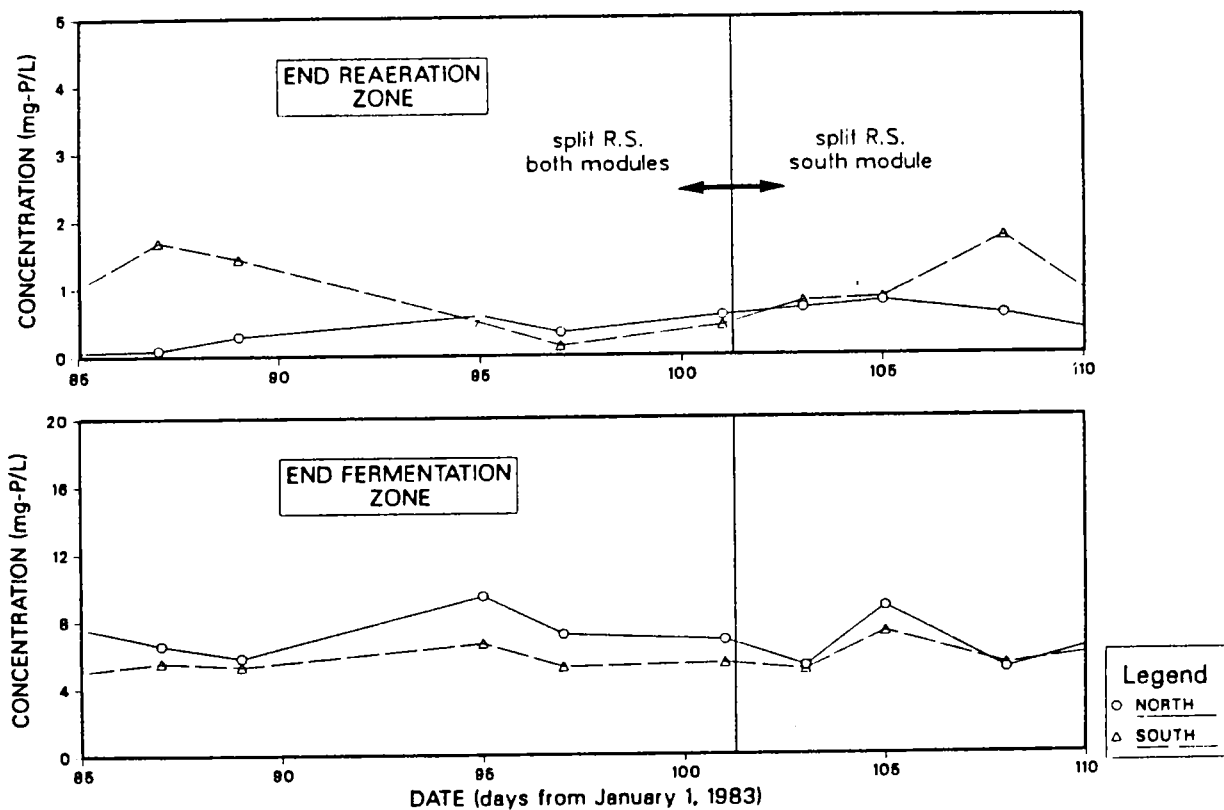Fig. 8 - Return sludge split test #2:nitrate-nitrogen.

Fig. 9 - Return sludge split test #2:orthophosphate.

TABLE 6 EFFECT OF RETURN SLUDGE SPLITTING ON PHOSPHORUS RELEASE

| | Test 1 | | Test 2 | |
	North Module	South Module	North Module	South Module
Return Sludge Entry	split	fermentation	fermentation	split
Ferm. ortho-P (mg/L)	5.5	4.0	7.0	6.5
Flow through Ferm. (L/s)	90	120	123	93
P-release (mg/L of RS)	7.0	4.0	10.4	12.2

It can be concluded from these data that the use of a split return sludge mode of operation does indeed give rise to an enhanced P release from the micro-organisms that enter the fermentation zone; however, this enhanced release from a reduced mass of micro-organisms only marginally affects the overall removal of P in the total bioreactor. For instance, data shown on Fig. 7 for orthophosphate in the reaeration zone indicate that the north module provides a slightly better and more consistent effluent between day 17 and day 38 ($\bar{x}$ = 0.10 mg-P/L; s.d. = 0.12 mg-P/L) than does the south module ($\bar{x}$ = 0.46 mg-P/L; s.d. = 0.36 mg-P/L). The data points for day 12 were not used in this comparison because they were gathered within one hydraulic retention time of the process change. It is possible that the positive effect of the split return sludge mode of operation would be greater if the facility were more highly loaded because the unit rate of phosphorus uptake in the aerobic zone would then be more important.

Nitrate results within the bioreactor show an increased nitrate content at the end of the first anoxic zone of the split return sludge module during Test 1 (see Fig. 6), but the disparity has been corrected by the time Cell 1F and Cell 1A have been reached (see Fig. 5). This indicates that pre-denitrification has been somewhat retarded in the split return sludge mode, but that the post-denitrification zone (second anoxic zone) is capable of handling the extra nitrates under the tested loading conditions. During Test 2 (Fig. 8) the same reduction in pre-denitrification capacity did not occur in the split return sludge module. This occurrence would seem to indicate that more simple substrate was passing through the fermentation zone to the pre-denitrification zone to quickly denitrify the nitrates entering that zone via the return sludge flow which by-passed the fermentation zone.

Effect of Primary Sludge Thickener Supernatant

The independent feeding of primary sludge thickener supernatant to each module of the bioreactor was tested on two occasions during 1983. General plant operating data during these test periods are summarized in Table 7. A slight difference in measured MLSS in Tests 3 and 4 is explainable in terms of the load entering the plant, caused by variations in both raw sewage load and thickener supernatant strength. During the periods of time when thickener supernatant was being fed to one module only, that module received approximately 45 percent of the incoming sewage flow plus the thickener supernatant, while the other module received 55 percent of the incoming sewage flow. This was necessary in order to maintain equal hydraulic loading

TABLE 7   PLANT OPERATING CONDITIONS DURING THICKENER SUPERNATANT TESTS

	Test 3	Test 4
Dates	Aug.-Sept. 1983	Sept.-Oct. 1983
SRT(d)	22	23
MLSS (mg/L)(avg.)	1550	1200
Mixed Liquor T (°C)	20-21	16.5-18.5
Inf. TKN/COD (avg.)	n/a	0.12
Inf. Q (L/s)(avg.)	150	135
Return Sludge Flow (L/s)	120	120
Thick. supernatant VFA (mg/L as HAc)(avg.)	140	110
Thick. supernatant Q (L/s)(avg.)	11	11

rates, because the thickener supernatant flow rate approaches 10% of the plant inflow.

The quality of the bioreactor effluent, in terms of both nitrate-nitrogen and dissolved orthophosphate, was significantly affected by the presence of thickener supernatant. The plots titled "end reaeration zone" on Figs. 10 and 11 show that both nitrates and ortho-P increased dramatically
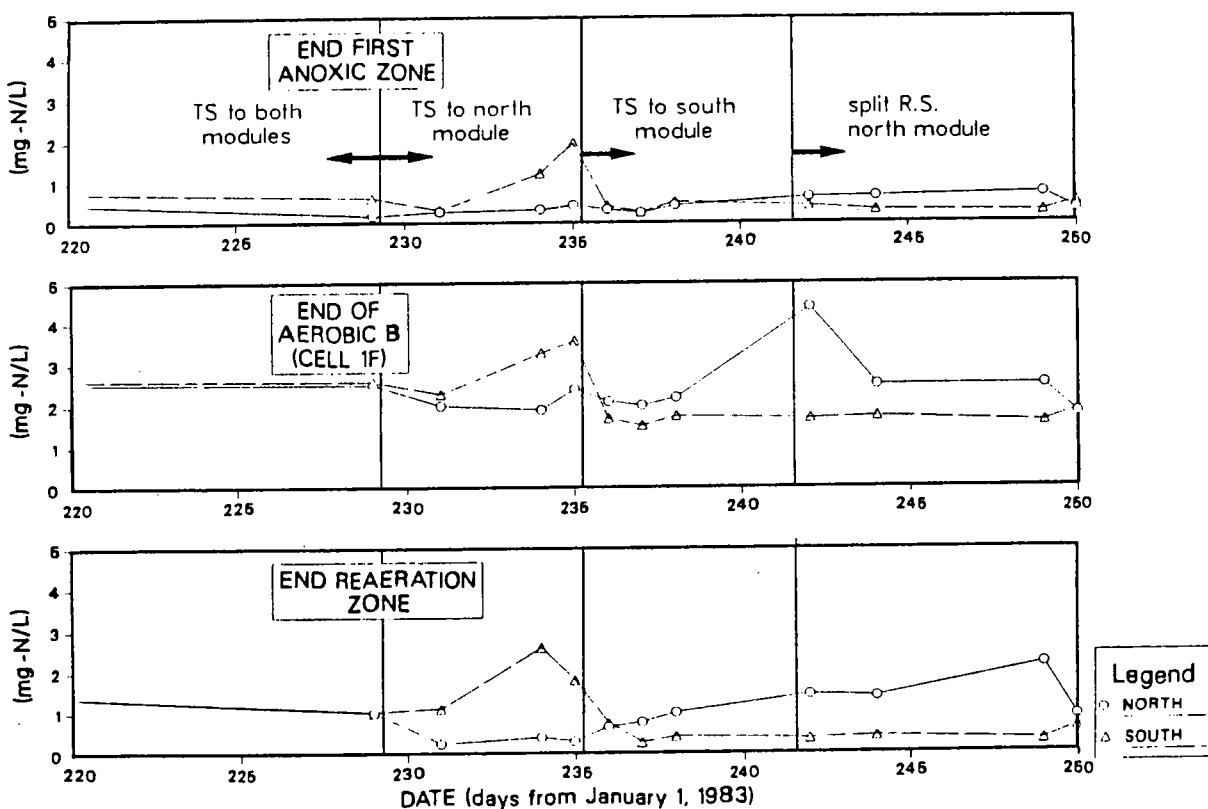
Fig. 10 - Thickener supernatant (TS) test #3:nitrate-nitrogen.
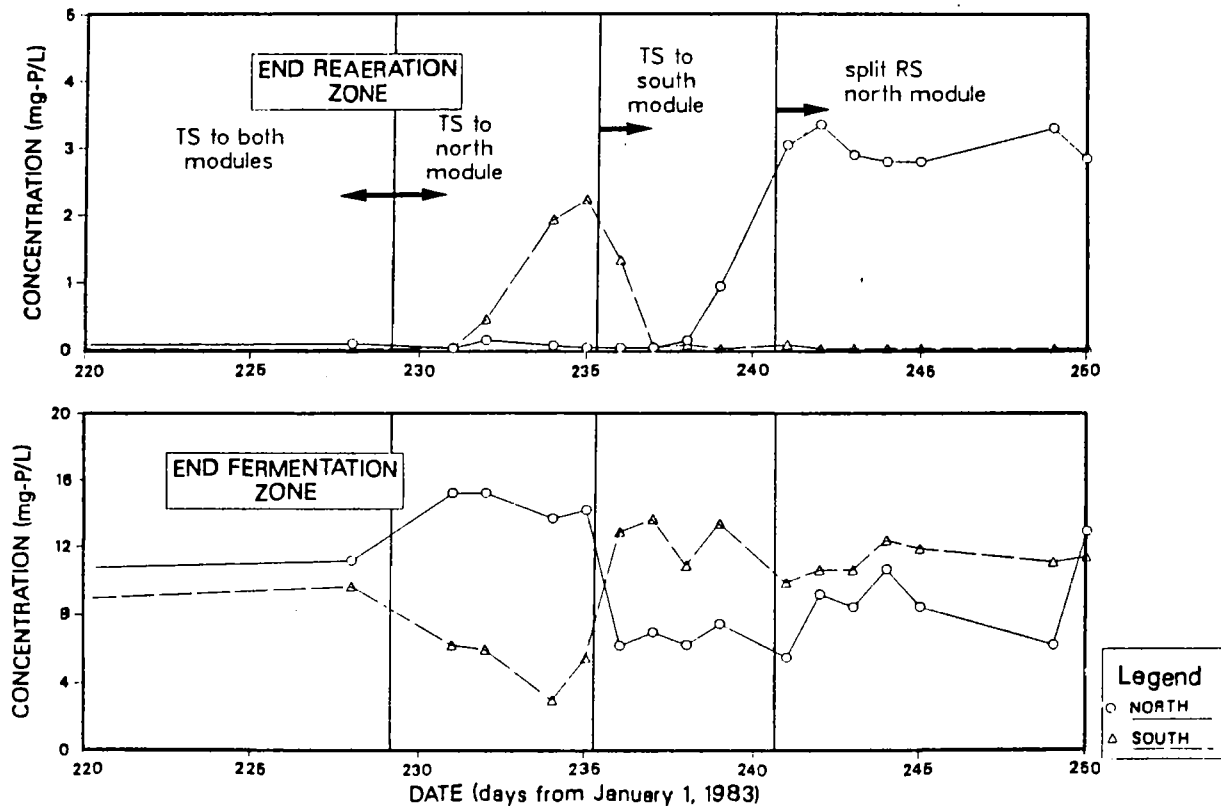
Fig. 11 - Thickener supernatant (TS) test #3:orthophosphate.

within a few days of the normal thickener supernatant addition being removed from either module. Conversely, the subsequent addition of thickener super-natant to the deprived module brought about a sharp improvement in process efficiency. Fig. 12 indicates that P removal in the North module was low under the initial condition where the thickener supernatant was split between both modules. There was subsequently no significant further deterioration when the supernatant was all transferred to the South module. However, when the total supernatant flow was changed to the North module, its phosphorus removal efficiency was markedly improved in three days, while the South module effluent deteriorated to some extent during the same time period.

These results indicate strongly that the micro-organisms necessary to uptake large quantities of phosphorus were continuously present in both modules, but that the substrate conditions required to stimulate the phos-phorus release and uptake mechanisms were not present when thickener super-natant was denied. Whether the necessary micro-organisms for good P uptake would remain in the biomass if it did not receive thickener supernatant for long periods of time is an unanswered question which may be very important for maintaining the prerequisite biomass mixture for good P removal in facilities treating fresh sewage, or sewage with a low ratio of BOD:P.

Reference to Fig. 11 suggests that it may be valid to consider that P removal organisms have a "memory" for their task without being subjected to the appropriate conditions in the fermentation zone of the reactor on every return cycle. After thickener supernatant was removed from the South module on Day 228 and subsequently from the North module on Day 235, in both cases
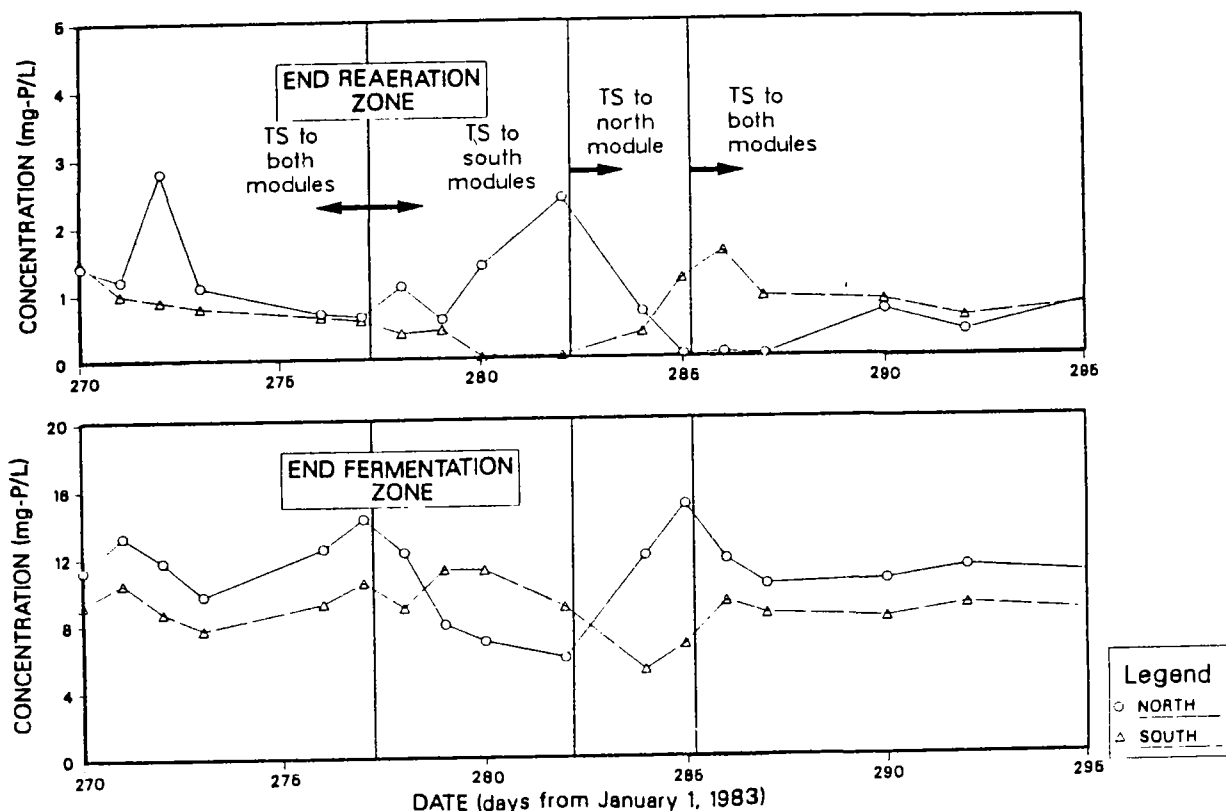
Fig. 12 - Thickener supernatant (TS) test #4:orthophosphate.

it took some 4 to 5 days before good P removal was lost, even though the nominal hydraulic retention time (NHRT) was less than 2 days. However, an equally plausible explanation can be advanced that takes into account the lack of a plug-flow process. Due to the potential for back-mixing between cells in the bioreactor and the presence of a large internal recycle flow, it is to be expected that a retention time experiment would show substantial variance and skewness in its data plot. Considerably more than one NHRT would be necessary to cause a large change in effluent quality under such conditions.

The effects of thickener supernatant on the internal conditions within the bioreactor are also significant. Fig. 10, for instance, shows that nitrates at the end of the pre-denitrification zone can rise sharply in that module which is denied thickener supernatant. Such a result indicates quite strongly that the thickener supernatant plays a significant role in maintaining a high rate of denitrification in the first anoxic zone. By reference to Fig. 10, the subsequent effect of this loss of denitrification can be traced through the rest of the bioreactor. The long term result of such an occurrence will be the recycling of extra nitrates in the return sludge to the fermentation zone which can have a further detrimental effect on the P release therein, as well as a potentially compounding effect on the increasing nitrate levels in the pre-denitrification zone. This longer term impact is indicated by the lower graph on Fig. 11, which shows the release of P in the South module fermentation zone to deteriorate over a six or seven day period, even though the loss of thickener supernatant would be completely felt in that zone within a few hours.

   With further reference to Fig. 11, it is interesting to note that the splitting of the return sludge entry to the bioreactor on Day 241 was successful in increasing the P release that occurred in the fermentation zone (both in terms of measured concentration and mass per unit volume of return sludge), but it did not improve the situation with respect to the dissolved ortho-P leaving the bioreactor. It is possible, however, that it did prevent the effluent ortho-P from continuing to an even higher level than shown on Fig. 11.

Maximization of VFA Production in Thickener

   In November, 1983 it was decided to increase the effectiveness of VFA production in the primary sludge thickener by reducing the rate at which thickened sludge was wasted from the bottom of the thickener. This had the effect of increasing the solids content of the wasted sludge and increasing the sludge blanket depth in the thickener (see Fig. 13). The added SRT in
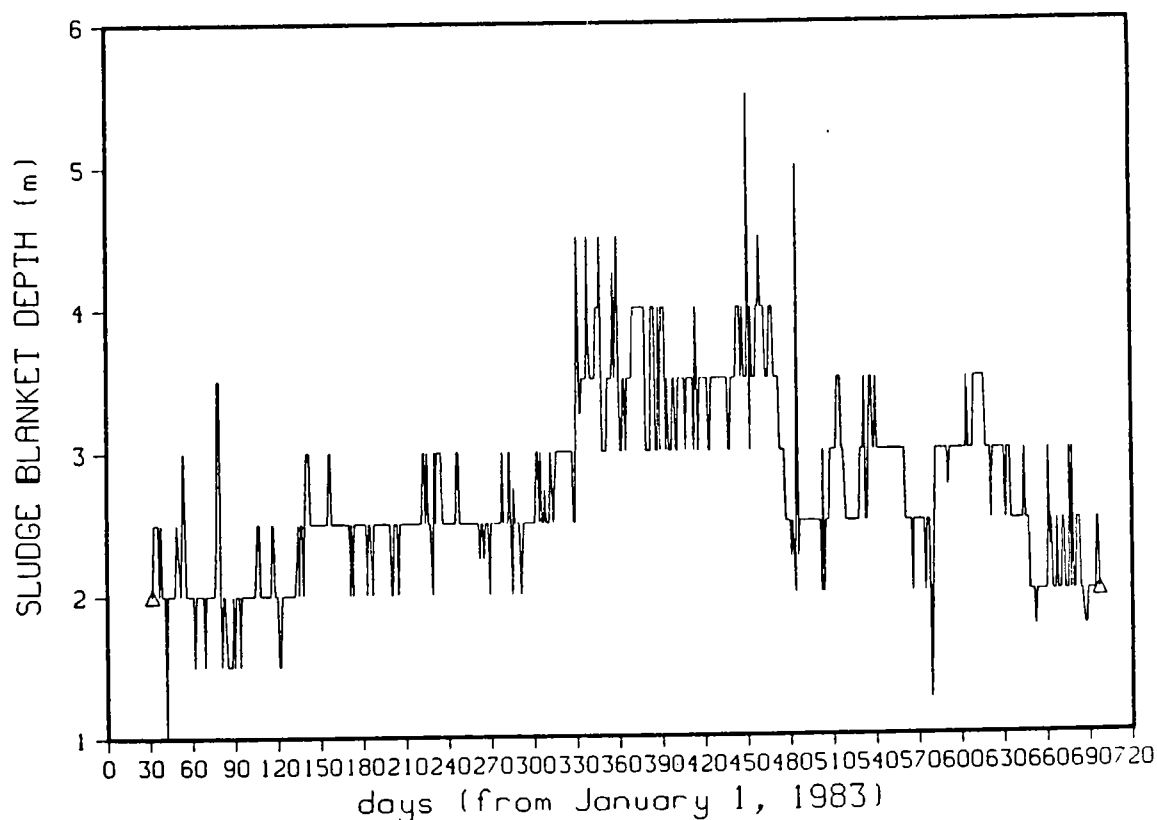
Fig. 13 - Thickener sludge blanket depth.

the thickener promoted greater anaerobic degradation of the complex organics
in the primary sludge, with the result that the thickener supernatant soon
responded with an increased VFA concentration (see Fig. 14). To speculate on
the effects that thickener operation has on phosphorus release and subsequent
uptake in the bioreactor, it is necessary to look at the time plots of both
anaerobic zone ortho-P (cell 2A on Fig. 5) and re-aeration zone ortho-P (cell
1A on Fig. 5). These time plots are shown on Figs. 15 and 16, respectively.

A general comparison of Figs. 14 and 15 shows that there is some degree
of direct correlation between VFA production in the thickener, and phosphorus
release in the anaerobic zone of the bioreactor. However, the increased P
release is not well-correlated with a reduced P content in the re-aeration
zone, except for one thirty-day period of time just prior to day 600. A
somewhat better correlation apparently exists between thickener sludge
blanket depth (Fig. 13) and re-aeration zone ortho-P (Fig. 16).

Obviously, it is not useful to simply maximize the VFA production from
primary sludge. Such maximization appears to improve P release in the
anaerobic zone of the bioreactor, but not to necessarily improve the
subsequent P uptake in the aerobic zone of the bioreactor. The very low
ortho-P results obtained in the re-aeration zone prior to day 600 appear to
be due to a combination of high VFA concentration in the thickener
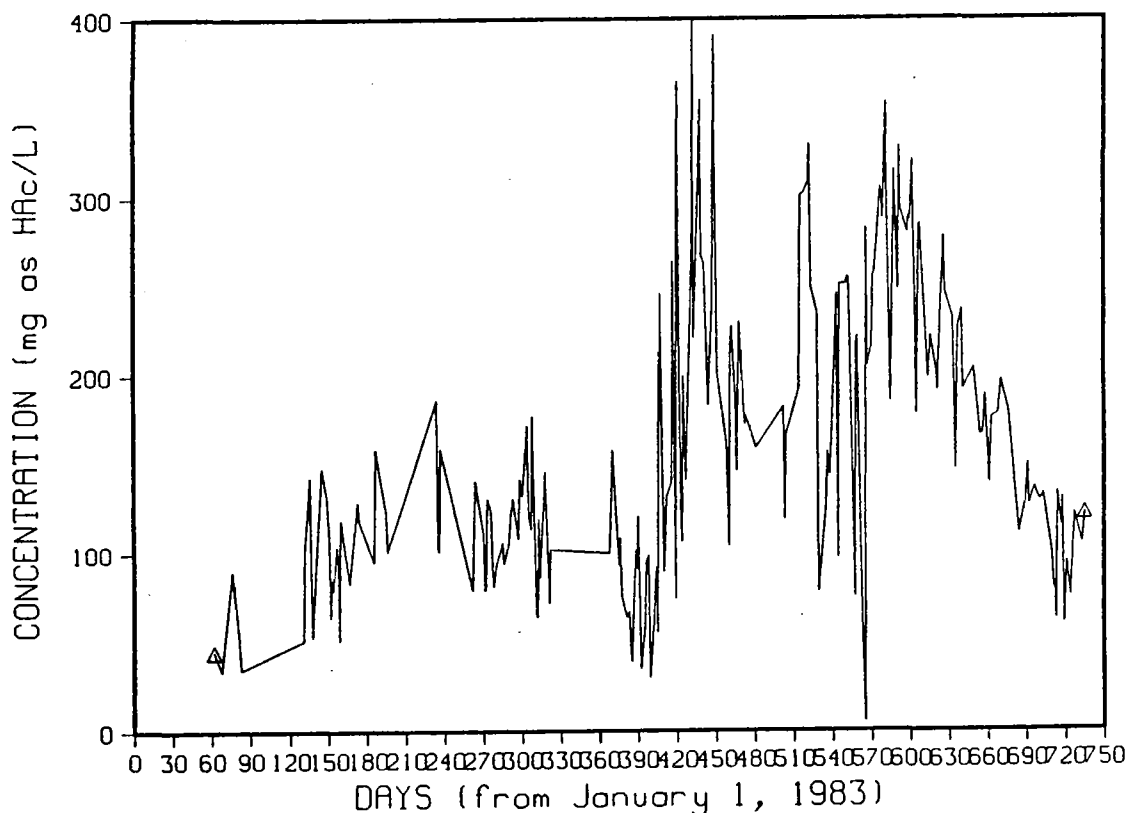supernatant, along with a low thickener sludge blanket depth.

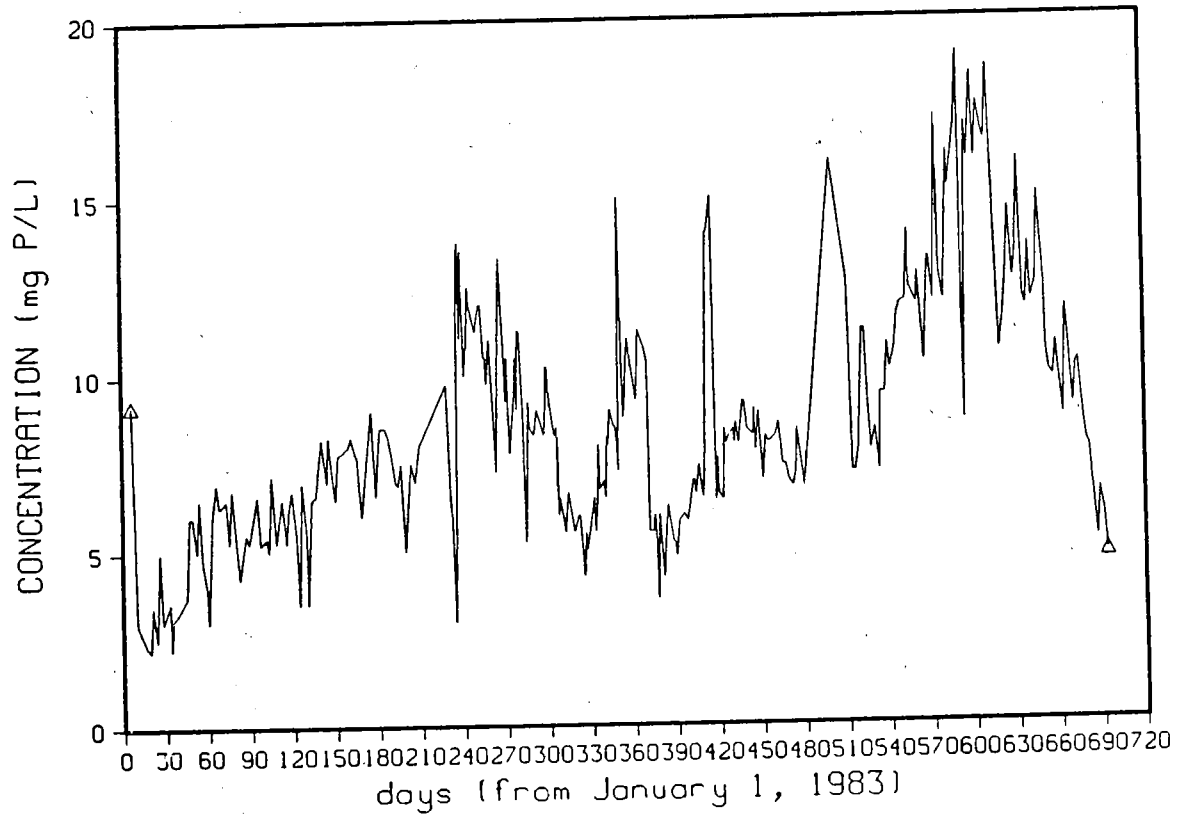Fig. 14 - Thickener supernatent VFA concentration.

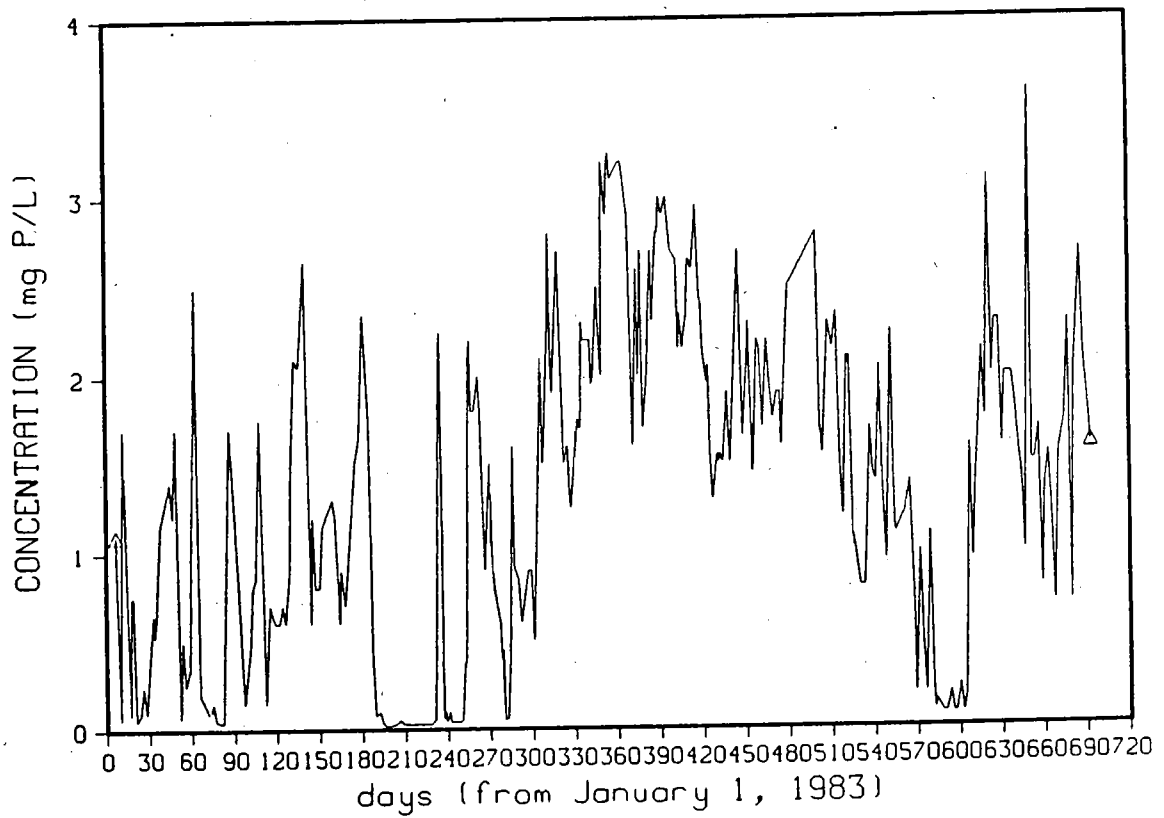Fig. 15 — Otho-P in bioreactor fermentation zone.

Fig. 16 — Ortho-P in bioreactor re-aeration zone.

It is well documented from previous work that high concentrations of VFA's are beneficial to the P removal process, but it appears that if these VFA's are produced by a long anaerobic SRT in a primary sludge thickener (as evidenced by a deep sludge blanket), there are other products formed which inhibit at least the P uptake phenomenon in the aerobic zone of the bio-reactor. At this time, a prime candidate as the culprit is the sulphide ion, as some indication of its presence in the sludge thickener has been found at times of high sludge blanket thicknesses. Furthermore, the sulphide ion is known to be toxic to a variety of organisms.

CONCLUSIONS

1. General plant effluent quality over the two year period 1983-84 shows median values of 0.8 mg/L total-P, 0.77 mg/L ortho-P, 1.5 mg/L TKN, and 1.8 mg/L nitrate-N. The effluent P requirement of 2 mg/L has been met 90% of the time, while the total nitrogen requirement of 6 mg/L has been met considerably more than 90% of the time.

2. During that portion of the two years when the plant was operating in a single-module mode, the load on the facility was at or greater than the design value. Effluent results during that time showed medial values of 1.1 mg/L for ortho-P (as opposed to 0.77 mg/L for the entire two years), 0.04 mg/L for ammonia-N, and 2.0 mg/L for nitrate-N. Compliance with the effluent quality requirements was statistically better than that achieved over the entire two year period. There is, therefore, little doubt that the plant will be able to handle at least the design flow of 23 ML/d.

3. The operation of the Kelowna facility in a split return sludge mode (50% to the fermentation zone, and 50% to the pre-denitrification zone) does increase the release of P in the fermentation zone, but has only slight overall impact on P removal, at least under the low loading conditions that prevailed at the Kelowna plant during 1983. The effect on nitrogen forms throughout the process was minimal, with the exception of some increased nitrate in the pre-denitrification zone, which was subsequently removed in the post-denitrification zone.

4. The presence or absence of primary sludge thickener supernatant in the fermentation zone of the bioreactor had a marked influence on both the P and N profiles thorughout the bioreactor. Overall P removal efficiency quickly dropped to levels found in conventional activated sludge plants when the thickener supernatant flow was removed. Conversely, in the situation where one of the reactor modules was marginally achieving good P removal with one-half of the thickener supernatant, the doubling of the supernatant input quickly restored excellent P removal. When the thickener supernatant flow was changed from one module to the other, P removal capabilities were quickly lost in the one module, and even more quickly restored in the other. It can be concluded that the short-chain organics (VFA's) provided by the primary sludge thickener are an important additive to the process, but that their absence for some period of time (at least 6 days) does not result in a die-off of the P removal organisms in the mixed liquor.

5. Increasing the production of volatile fatty acids from primary sludge under anaerobic conditions does not appear to automatically improve P-removal efficiency in the bioreactor. There is some suspicion that compounds detrimental to the P-removal process may also be formed if the anaerobic detention time of the primary sludge is extended too far.

ACKNOWLEDGEMENTS

REFERENCES

Barnard, J.L. (1982). The influence of nitrogen on phosphorus removal in activated sludge plants. Wat. Sci. Tech., 14, 1/2, 31-45.

Comeau, Y. (1984). Biochemical Models for Biological Excess Phosphorus Removal from Wastewater. M.A.Sc. Thesis, Dept. of Civil Engineering, University of British Columbia.

Davelaar, D., Davies, T.R. and Wiechers, S.G. (1978). The significance of an anaerobic zone for the biological removal of phosphate from wastewaters. Water SA, 4, 54-60.

Ekama, G.A., Siebritz, I.P. and Marais, G.v.R. (1983). Considerations in the process design of nutrient removal activated sludge processes. Wat. Sci. Tech., 15, 3/4, 283-318.

Ekama, G.A., Marais, G.v.R. and Siebritz, I.P. (1984). Biological excess phosphorus removal. Chapt. 7. Theory, Design and Operation of Nutrient Removal Activated Sludge Processes. Water Research Commission, Pretoria, South Africa.

Fuhs, G.W. and Chen, M. (1975). Microbiological basis of phosphate removal in the activated sludge process. Micr. Ecol., 2, 119-138.

Nicholls, H.A. and Osborn, D.W. (1979). Bacterial stress: prerequisite for biological removal of phosphorus. Jour. Wat. Poll. Cont. Feb., 51, 557-569.

Oldham, W.K. and Dew, H.P. (1979). Biological nitrogen and phosphorus removal at low temperatures. Presented at Annual Conf., Wat. Poll. Cont. Fed., Atlanta, Georgia.

Oldham, W.K. and Stevens, G.M. (1984). Initial operating experiences of a nutrient removal process (Modified Bardenpho) at Kelowna, B.C. Can. Jour. of Civil Eng., 11, 474-479.

Siebritz, I.P., Ekama, G.A. and Marais, G.v.R. (1983). A parametric model for biological excess phosphorus removal. Wat. Sci. Tech., 15, 3/4, 127-152.