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SEMINAR NOTES

NUTRIENT CONTROL TECHNOLOGY

SPONSORED BY

ALBERTA ENVIRONMENT

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FEBRUARY 7-8, 1980

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EUTROPHICATION: CONSIDERATIONS FOR NUTRIENT REMOVAL

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Presented at

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CANADA

Eutrophication by D.R.S. Lean N.W.R.I.
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Environmental problems will be understood better when we realize that what we call pollution is the result of processes, which exceed the capacity of the biogeochemical processes in nature to respond to societies'wastes. Industrial man cannot expect to live in a pristine environment but what seems surprising to me is not how bad the environment is but how good it is. Clearly, nature has a considerable, but not infinite, capacity to forgive. By learning more about biogeochemical processes we may be able to adjust the processes imposed by man's activities to be more compatible with nature.

A consequence of the structure of our industrial society is that most people are displaced from the raw materials that they need (food, minerals, energy). This linear flow results in resource depletion and waste accumulation.

The influence of nutrients on the productivity of lakes has been recognized for many years. For centuries, especially in Europe and Asia, people wishing to enhance the fish harvest in ponds would add fertilizer. More recently, controlled experiments have been done on the addition of specific nutrients to a wide variety of lake types and lake enclosures. Phosphorus has been identified as the nutrient element which usually causes the most significant increase in algal biomass but combinations of the appropriate form of phosphorus and some nitrogen compounds stimulate an even greater response.

Rigler (1975) gave an invited lecture to the Congress of Theoretical and Applied Limnology. His philosophical theories may discourage some readers but he suggested that rather than being totally absorbed in the endless number of fascinating aspects one discovers in lakes, we should select factors which allow us to become predictive. He emphasized the work of Dr. R.A. Vollenweider as an example of the type of predictive limnology he thought was most useful to managers. This type of research Rigler termed holistic, a study of the intact system rather than a small or isolated part (reductionist). He outlined a series of experiments which he thought would be more productive. Following Sakamoto, (1966) Dillion and Rigler (1974a) showed that total phosphorus at spring overturn was highly correlated with the mean chlorophyll a concentration in summer. Using Vollenweiders approach Dillon and Rigler (1974b) were able to predict (P_{spr}) using the formula $\frac{J\ (1-R)}{Q}$ where J is the total annual input from all para limnetic sources, Q is the annual water discharge, and R is the phosphorus retention, i.e. the fraction of J left in the lake. R is equal to $\frac{J-Pout}{J}$. R is of critical importance to those constructing reservoirs. They may or may not wish to trap phosphorus within the new lakes.

Rigler (1975) went on to show how it is possible to predict J knowing geology and land use. This is discussed below. Predicting Q and R means that we can predict how green a lake will be without even leaving the laboratory.

$$\log_{10} [Chlora] = 1.53 \log_{10} [P]_{spr} - 1.14 \qquad R = 0.96$$

Kirchner and Dillon (1975) have been able to predict R as a function.

of the hydraulic load Q/A. A is the lake surface area.

$$R = 0.426\, e^{-0.271\, Q/A} + 0.574\, e^{-.0095\, Q/A}$$

An integral P input is aeolian. This may represent the principal P input to lakes with small watersheds in igneous regions. Shore based measurements may overestimate P input and the poor data base is illustrated in the review of Lee et al (1979). Perhaps these data exist in the computers of our own government agencies but we have not reduced the values to a form to be used in our empirical equations.

I have been in the fortunate position of working with Dr. Rigler (1967-72) and being an associate of Dr. Vollenweider until the present. Rigler dismissed his early work as reductionist and not useful in predicting anything about lakes. Vollenweider considered his models were far too simple and thought Rigler's isotope work would eventually lead to useful information.

In 1968, Vollenweider produced a report "The Scientific Fundamentals of Eutrophication of Lakes and Flowing Waters, With Particular Reference to Nitrogen and Phosphorus as Factors in Eutrophication". Although based on a relatively scant data collection, this report set the stage for research over the past 11 years. Scientists throughout the world have been reeducated by his approach. The concepts of nutrient loading and lake response have become vital to responsible management of lakes.

This initial report was sponsored by the Organization for Economic Cooperation and Development (OECD), based in Paris, France. The OECD was formed in 1961 as an outgrowth of the Marshall Plan for economic recovery after World War II. In 1971, a large-scale cooperative research program on eutrophication monitoring was launched. Participants represented over 40 institutions in 18 countries throughout the world, including the United States, Japan and Australia. Participation was largely voluntary but, with Dr. Vollenweider's leadership, all agreed to use similar methods of analysis and monitoring. After 8 years, a final report is coming out next year.

I have read the draft report and Dr. Vollenweider very generously allowed me to show you some of his graphs.

The underlying concept is that a lake is an open system. The mass balance equation is applicable to any substance involved in lake dynamics.

$$\frac{\Delta M}{\Delta T} = I - O - (S-R)$$

Where ΔM = storage gain or loss in time Δt, I = external load,

O = outflow loss, S = loss to sediments, R = sediment return (internal loading) (S-R) is difficult to measure but can be obtained by difference. Some researchers have attempted to subdivide the lake but Vollenweider considers it to be a stirred reactor, and produces average values rather than modelling gradients etc. Furthermore, rather than predicting M over days or even weeks, Vollendweider considers T to be 1 year.

For a lake in steady-state $\Delta M/\Delta T = 0$. In reality no lake is in a steady state but oscillates around a mean except during periods when the loading is altered. I is the sum of all the inputs (flow Q times the concentration (M) and simplifications using yearly flow times average concentration are possible.

Vollenweider's most recent simplification of the mass balance equation is:

$\overline{P}$ = Predicted Annual mean concentration of total phosphorus ($mg.m^{-3}$)

$$= \frac{\overline{P}_i}{1 + T(W)}$$

where T(W) = water residence time (y)

$$= \frac{V\bar{o}}{Q_y} = \frac{\text{lake volume } (m^3)}{\text{outflow } (m^3, y^{-1})}$$

and $\overline{P}_i$ = annual mean inflow of total phosphorus ($mg.m^{-3}$)

Dillon and Rigler (1974) used a slightly different modification predicting $(P)_{spring}$. Vollenweider (per. com.) claims there is little difference between P_{spring} and $\overline{P}$.

From all the OECD data, Vollenweider provides the following correlations:

Fig 1. The relationship between predicted P from P loading and water residence time and measured P

Fig 2. Predicted P and mean annual chlorophyll in the euphotic zone

Fig 3. Mean annual chlorophyll and maximum chlorophyll; and

Fig 4. Mean measured phosphorus and mean annual chlorophyll.



Figure 1

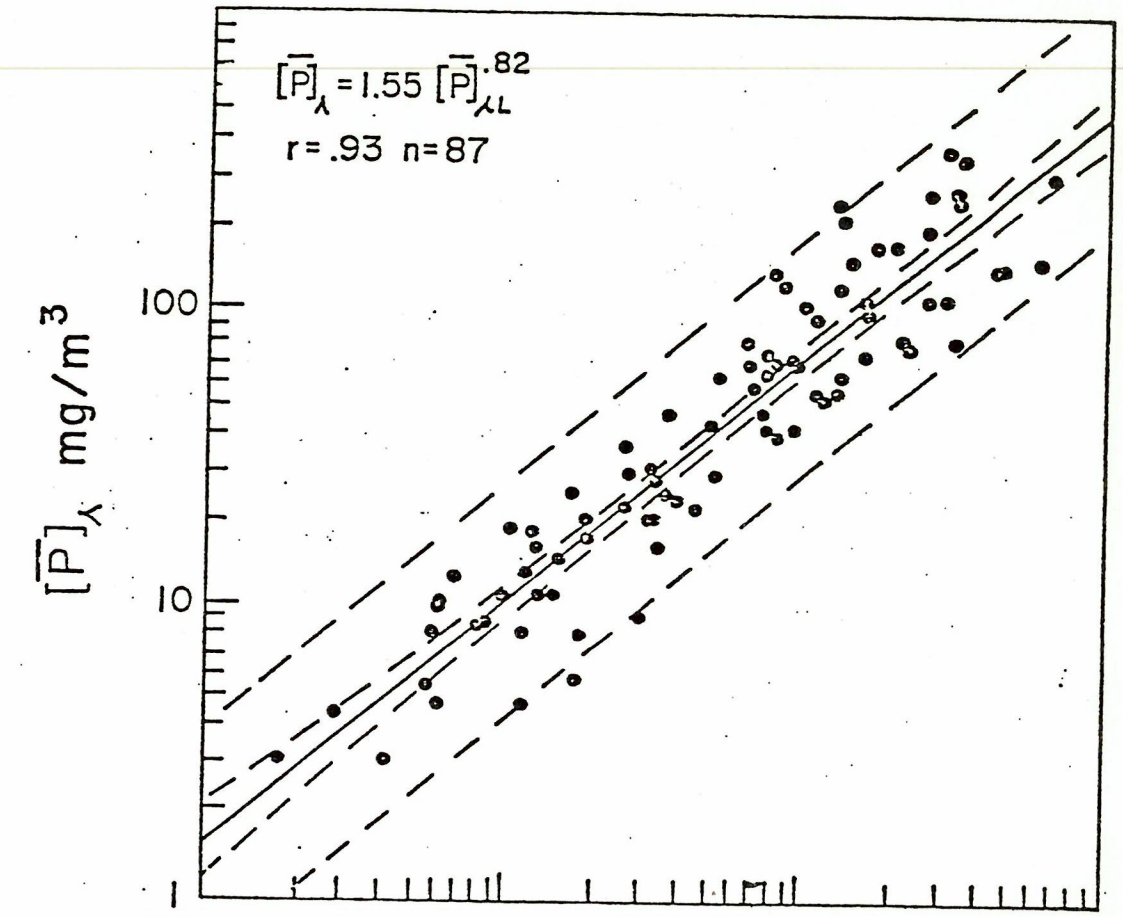


Figure 2

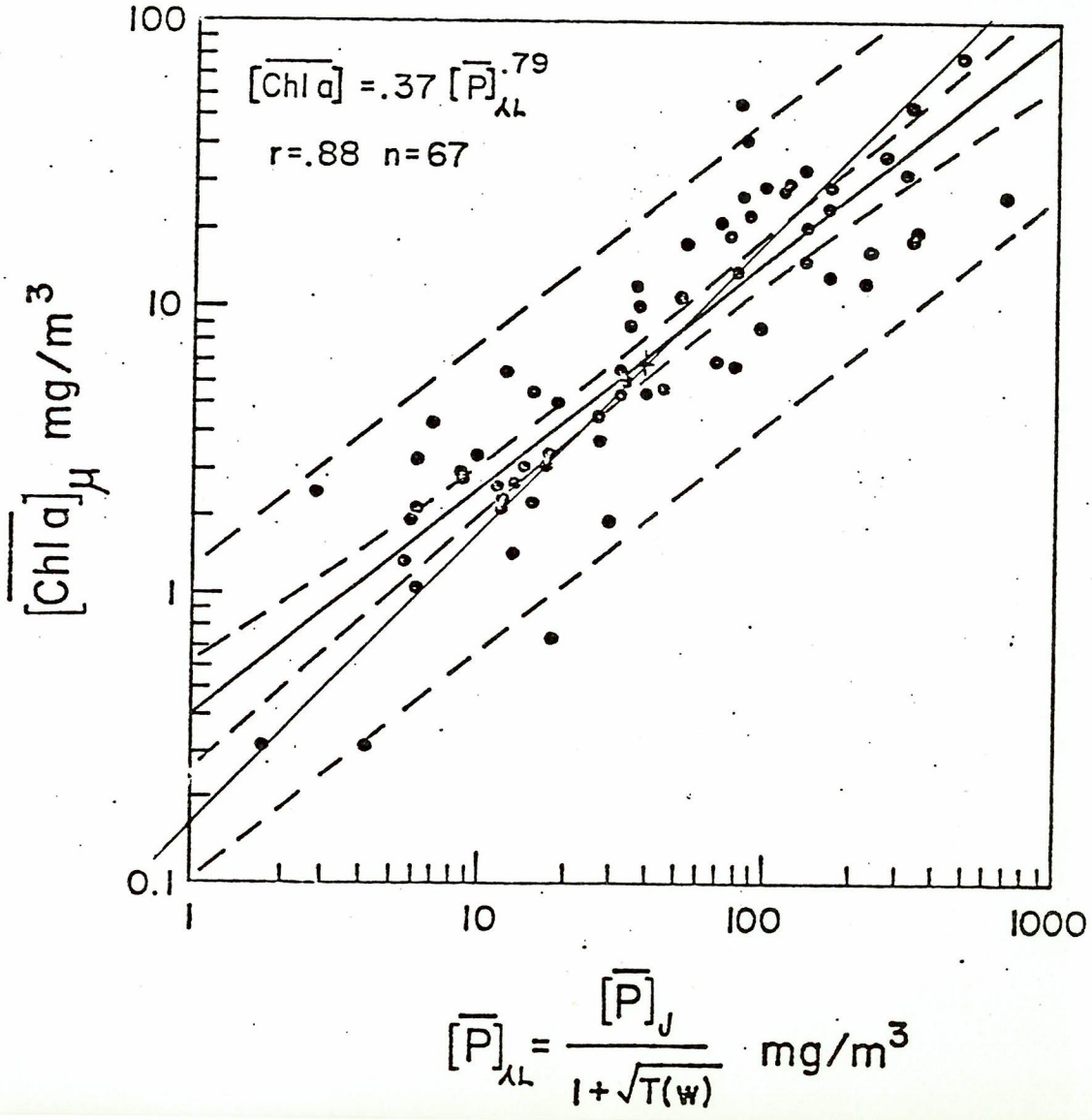






Figure 3

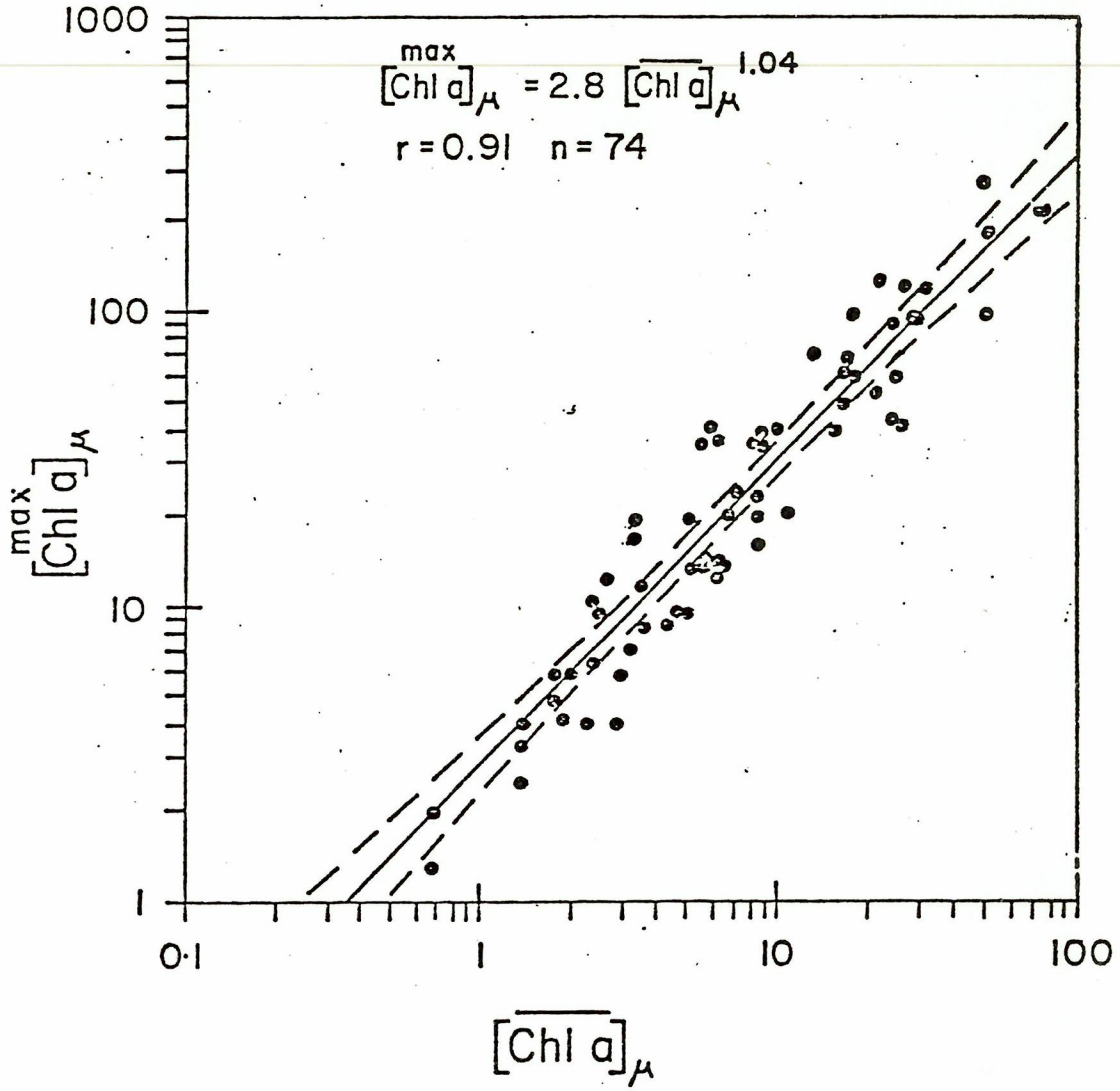
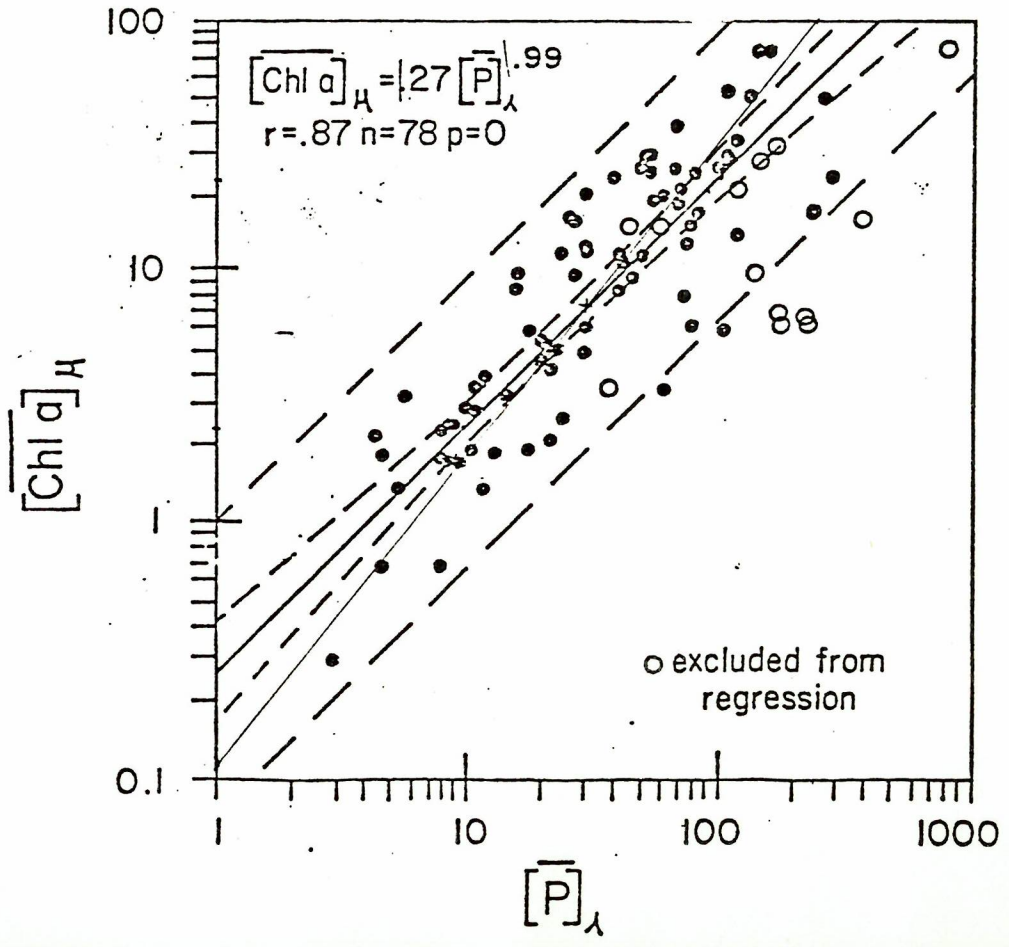


Figure 4



The relationships given above can also be used for diagnostic purposes. If for any particular lake the chlorophyll or $\bar{[P]}_s$ is higher than predicted it may mean the budget is bad or internal loading is significant. If lower, the sediments may be acting as a sink removing phosphorus or there is considerable attached rather than planktonic algae. If chlorophyll is not on the line as predicted the lake may not be P limited or toxic substances may be inhibiting the growth of algae.

This provides an explanation for the vital discoveries of Rawson (1955) and Ryder (1964) who thought that total dissolved solids (TDS) and mean depths were all that was required to predict phytoplankton biomass and fish yield. Phosphorus is an insignificant fraction of the total solids in lakewaters. However, the TDS may be a function of P export from the drainage basin. Mean depth was previously used in Vollenweider's equation but its influence is now included in the water residence time function. The water in a shallow lake has shorter residence time than in a deep lake.

But, where do the nutrients come from? In 1978, the International Reference Group on Great Lakes Pollution from Land Use Activities (PLUARG) made its final report to the International Joint Commission. Diffuse loadings of phosphorus from the land were shown to account for about 30-50 percent of total inputs, but less than 35 percent of this was considered biologically available. Most of the rest for Lakes Erie, Ontario and Michigan comes from municipal treatment plants, urban drainage, or in the cases of Lakes Erie and Ontario from the upstream Lake. Atmospheric inputs become a significant portion of the loading for Lake Huron, Michigan and Superior because of sparse urbanization.

The major inputs to Superior are from forest runoff and diffuse loadings from the atmosphere, whereas, for Lakes Ontario, Erie and Michigan, the inputs are from urban activities and farming. PLUARG, perhaps wisely, did not produce calculations of the average phosphorus export for each land use. When this is done, a generalization emerges which may be useful. Rates of 0.1, 1, and 10 kg/ha/yr are lost from forest, farm and urban society respectively (rates times 100 give $mg/m^2/yr$.) Clearly, establishing forests between farms and streams can improve the situation with respect to diffuse loadings from erosion. This must go along with improvements in sewage plants for phosphorus removal.

Dillon (1974), Dillon and Kirchner (1975) studied P export from forested watersheds and those containing small farms. The export from forested, igneous and sedimentary regions was 4.7 + 4 and 12 + 5 $mg.m^{-2}y^{-1}$ respectively. Where the land was partly used for pasture the value increased from 12 to 29. Dillon and Kirchner also reviewed literature from areas in Europe and elsewhere. It should be our aim to develop simple formula to predict export as a function of land use, geology, and population density. As far as I know no reliable export data has been obtained for prairie regions.

In Sweden considerable effort is being placed on reducing the P export at the source. Small ponds can be constructed which will retain

phosphorus and denitrify incoming combined nitrogen. They can be attractive and no odours are produced even downstream from feedlots.

The reservoirs should be designed to fit into the particular local requirement. The usual design would aim at trapping nutrients. This may require periodic draining and cleaning. Sediments make excellent top soil for use in restoring land that is unproductive. Using empirical relationships we now know that they should be fairly deep and Q/A should be maximized. In other words the flushing time should be as long as possible. Reservoirs with a rapid flushing time, especially those that are turbid do not act as nutrient traps.

In Ontario, Dr. Don McQueen of York University was aerating the hypolimnion of a small Kettle lake north of Toronto when a pig farm was located adjacent to the inflow. He was still able to maintain water quality better than that prior to aeration. This technique looks promising for future lake restoration.

It should be remembered that having more nutrients is not necessarily bad. Recall that fish harvests can be enhanced by nutrient enrichments. However, the consequence of changing nutrient levels may be a change in the species composition to less desireable forms. Similarly, a consequent change in algal species composition may be of concern. For example, a floating mass of rotting blue-green algae on the surface is more objectionable to the eye than the same quantity (on an areal basis) distributed throughout the water column. Furthermore, large amounts of attached algae, like Cladophora, may break off along a shoreline near a city and be swept onto a beach where rotting occurs with its consequent odours. In this situation, the water may be fine but the beach is foul. Avoiding such problems requires new information from scientists on details of the mechanisms of phosphorus and nitrogen uptake and the dynamics of lake metabolism. Where the nutrients go and the nature of the factors which promote the growth of some species over another are questions still not answered. The loading models of Vollenweider lead directly into studies of mechanisms.

Just as cellular biochemistry and studies on human metabolism have provided valuable insight into the processes which are important to human medicine, so it is vital to know more about the cellular biochemistry of lake organisms and the basics of lake metabolism. The uptake and release of phosphorus, nitrogen and carbon in lakes and the transformations of compounds containing them with respect to specific organisms are being measured using radioisotopes for phosphorus and carbon and the stable isotope of nitrogen. This permits new knowledge to be gained on the growth of specific algae in competition with bacteria and alternate fates of the algae resulting from decomposition and grazing by zooplankton.

This is of course completely neglecting the warnings of Rigler and we are now becoming reductionists. It is necessary if we are to answer specific questions eg. How does nitrogen influence algal biomass?

My understanding of the role of nitrogen in eutrophication stems from experiments in enclosures, some of which is published but most is not. NH_4 or NO_3 enrichment by itself shows little or negative chlorophyll enhancement over the control enclosures receiving no nutrient enrichment. Denitrification is the obvious explanation. P loading by itself often brings on blue-green algae which can supply most of the nitrogen required but this may result in an iron limitation. P loading by itself rarely achieves the same biomass as P + N together. This implies N-fixation cannot keep up to algal requirements. In lakes there is a third pathway involved which has been largely overlooked. N inputs to lakes are usually poorly monitored because nitrate moves so rapidly through groundwaters. Excess NO_3 is quickly denitrified. Phosphate enrichments cause phytoplankton to compete with the denitrifyers making the recycling of N far more efficient. The N/P ratio in the lake is then maintained as a simple function of chlorophyll a.

Only a small fraction of the organic matter which is produced in the surface waters sinks to the deeper waters. Growth and decomposition are almost in balance. Each atom of phosphorus is reused several thousand times before it is eventually and irreversibly locked up in the bottom sediments. Zooplankton filter the equivalent of about 20% of the entire volume of the lake epilimnion each day. The fraction which eventually falls to the deeper water contributes to oxygen depletion of the hypolimnion.

The form of the material settling out of lakes is difficult to describe. Using classical limnological technology, little identifiable material is recorded. However, with the aid of the electron microscope, and associated technology transferred into limnology from the biomedical sciences, it was shown recently that much of it is clumps of tiny fibrils which are produced at the cell surface of some algae and natural lake bacteria. These fibrils appear resistant to degradation, they contain chemical groupings known to bind heavy metals and other non-degraded toxic substances and they can aggregate to settle out. The distribution of toxic substances in the sediment correlates with fine grain, organic rich particles. They may act to accumulate toxic substances or concentrate toxic substances to be transferred to the benthos. Furthermore, this material is probably closely related to sewage sludge. A complete morphological examination is provided by Massalski and Leppard (1979), Leppard et al (1978).

Fibrils must be resistent to degradation so it is not surprising that we found this to be the predominate form of material settling out of lakes. This contradicts the conventional wisdom that suggests the

materials settling out are dead and living algae, bacteria and zooplankton fecal pellets. The material settling out may contain some easily oxidizible material but much of it is degraded only over a very long time.

In the 1960's the word eutrophication was popularized, based almost entirely on the idea that Lake Erie was dying. We were told that algae were growing so luxuriantly that upon dying and sinking, their remains used up the oxygen of the bottom waters. Recent reanalysis, however, shows that no trend is significant over the last 50 years, (Charlton, 1979, 1980). Furthermore, it is now possible to interpret, on a sound basis not only Lake Erie's oxygen depletion problem but oxygen depletion in lakes in general. The problems in the Central Basin of Lake Erie are a consequence of its shallow hypolimnion and not just the recent nutrient inputs. Just because a lake shows severe hypolimnetic oxygen depletion it does not necessarily mean it is suffering from recent eutrophication. It may be due to its morphometry.

Oxygen and organic matter are produced in the upper waters of lakes but some of the organic matter (less than 10% of the primary production) sinks into the dark, cold deeper zone where decomposition, and respiration results in oxygen consumption.

Thermal stratification restricts any replenishment of oxygen and the concentration decreases steadily until turnover. At this time lakes will reach equilibrium with atmospheric O_2 if the mixing occurs long enough. Sometimes, however, this does not happen.

Oxygen depletion can stress fish populations living in the deeper waters, can cause production of methane and hydrogen sulphide, and if sufficiently severe can trigger the internal loading of phosphorus. The latter effect can intensify summer phytoplankton blooms.

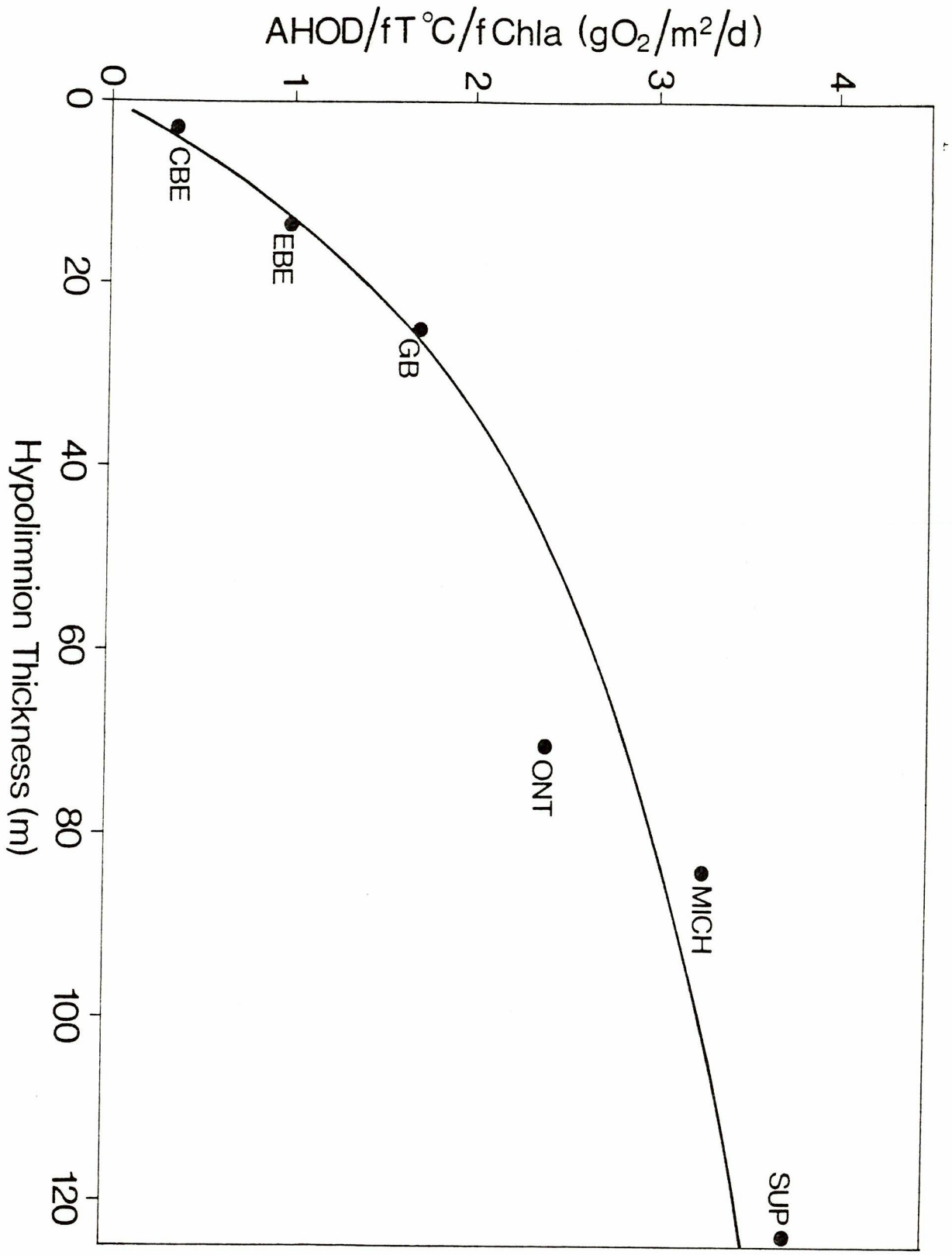
The total hypolimnion oxygen demand (THOD) of a lake is comprised of sediment oxygen demand (SOD) and water oxygen demand (WOD). Oxygen consumption is usually calculated as the mean loss in oxygen per litre volumetric hypolimnetic oxygen depletion rate (VHOD) which can be divided by the depth of the hypolimnion to give the areal hypolimnetic oxygen depletion rate (AHOD).

If AHOD's are influenced by productivity, Charlton thought that observed AHOD's from the Great Lakes could be normalized if the AHOD was divided by productivity and corrected for temperature. He used an expression derived from Vollenweider et. al. (1974)which relates productivity (P) to Chlorophyll a (c)

$$P = \frac{1.15\ C^{1.33}}{9 + 1.15\ C^{1.33}}$$

He also assumed a Q_{10} of 2 so that all observed AHOD's were corrected to a temperature of 4 C. The observed rate at 4 C would be

Fig. 5: Relationship between temperature and chlorophyll corrected AHODs and mean hypolimnion thickness in the Laurentian Great Lakes.



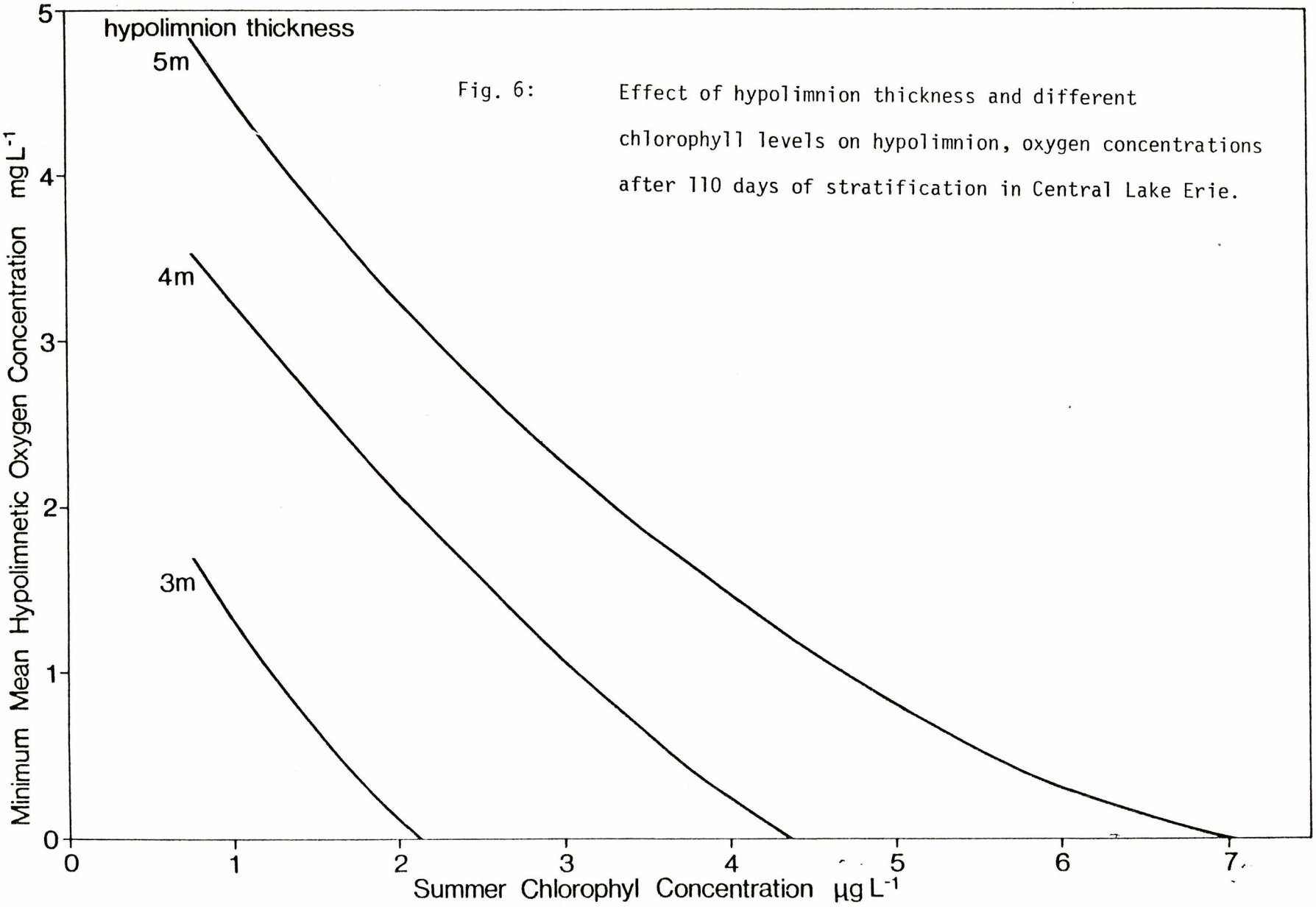


Fig. 6: Effect of hypolimnion thickness and different chlorophyll levels on hypolimnion, oxygen concentrations after 110 days of stratification in Central Lake Erie.

half that at 14 C. This correction is most important for Lake Erie's central basin, where the SOD and sediment storage are different from other lakes. He found the data (Fig 5) followed a rectangular hyperbola:

$$AHOD\ (corr) = AHOD\ max \frac{H_t}{K + H_t}$$

where K = 50

The upper limit shows no further change with increasing depth but the AHOD's are expressed per unit productivity. This means lakes must be deeper than 150 m before hypolimnetic oxygen depletion is a good indicator of productivity. If the data were not expressed per unit productivity the AHOD for Eastern Lake Erie is greater than for Lake Superior. When the productivity and temperature differences are factored out the relationship between H_t on AHOD is obvious. Chlorophyll and H_t each accounted for half the total variance. The curves (Fig 6) show the relationship between oxygen and cholorophyll at several hypolimnion depths.

Charlton reinforced his argument using data from Thienemann (1928) (Fig 7) and simply plotted it on a log-log scale. Using only mean depth of the lake a reasonable prediction of hypolimnetic depletion is obtained. It might be argued however that these data predate the eutrophication due to detergent phosphorus.

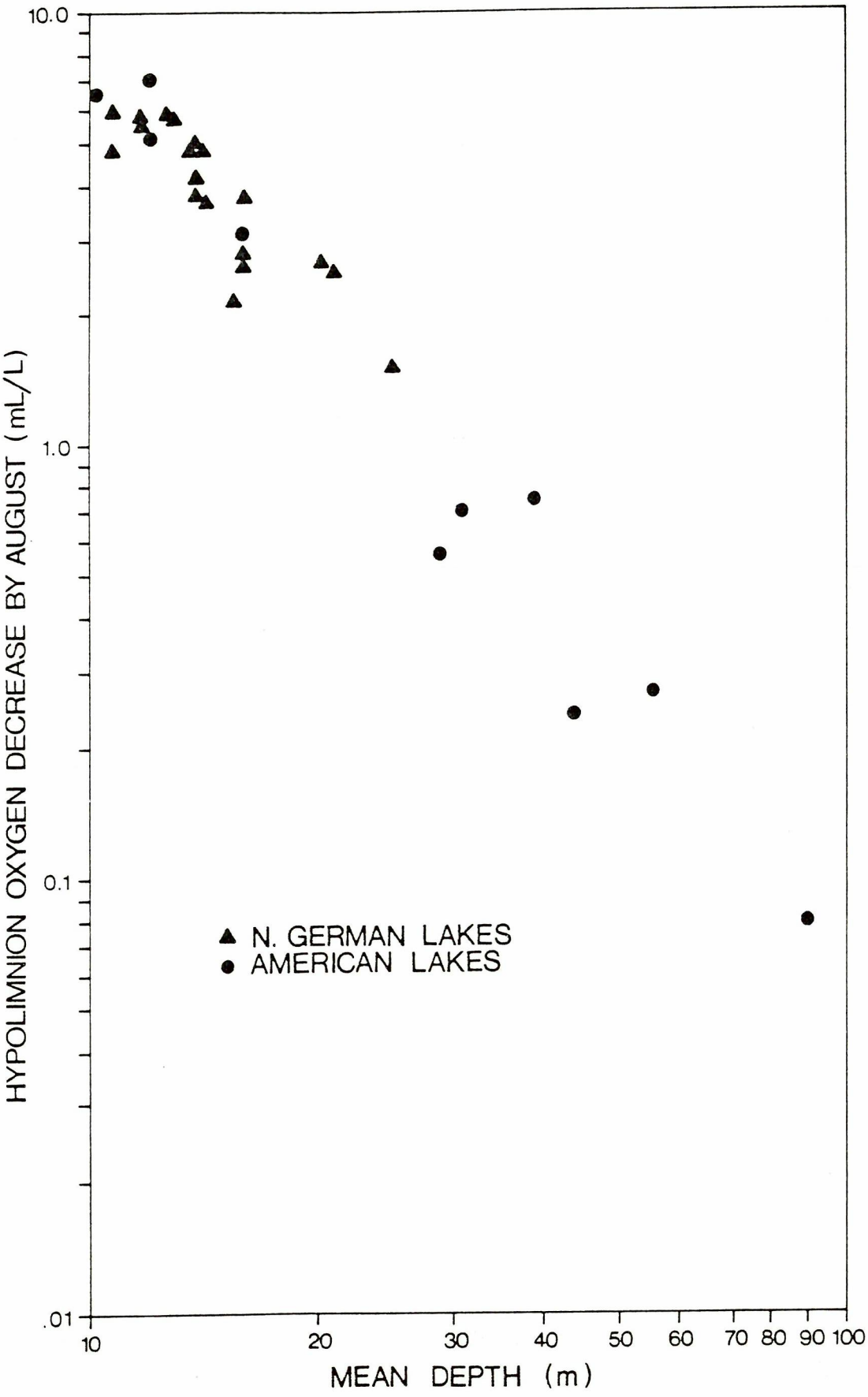
The decision for reducing P has certainly helped many small lakes and perhaps Lake Ontario. In Lake Erie, the improvement may have been in reducing Cladophora on the beaches and near shore chlorophyll blooms but these problems were poorly quantified. It has not significantly changed the hypolimnetic oxygen depletion problem.

Cornett and Rigler (1979) developed a predictive relationship for AHOD for several small lakes. It requires measuring or predicting phosphorus retention (R_p). The weakness in this model involves this prediction. It may be as difficult to obtain as actually measuring the O_2 directly. Charltons (in press) relationship uses only chlorophyll.

The areal hypolimnetic oxygen depletion (AHOD) was thought to be independent of lake depth and could be used as an indicator of organic productivity (Hutchinson 1937, 1957). He cautioned that it should be used only for lakes of mean depth of 20 to 75 m. This provision has been largely neglected, especially in discussions on Lake Erie. We know realize we must include a term for hypolimnion depth and temperature.

Charlton (1979, 1980) reexamined the data from 1929 to the present and found that the yearly fluctuations could be explained from differences in the thickness of the hypolimnion (2-5m) and the temperature of the hypolimnion (4-11 C). He reasoned that a shallow

Fig. 7: Relationship between oxygen depletion and mean depth in lakes studied by Thienemann (1928).



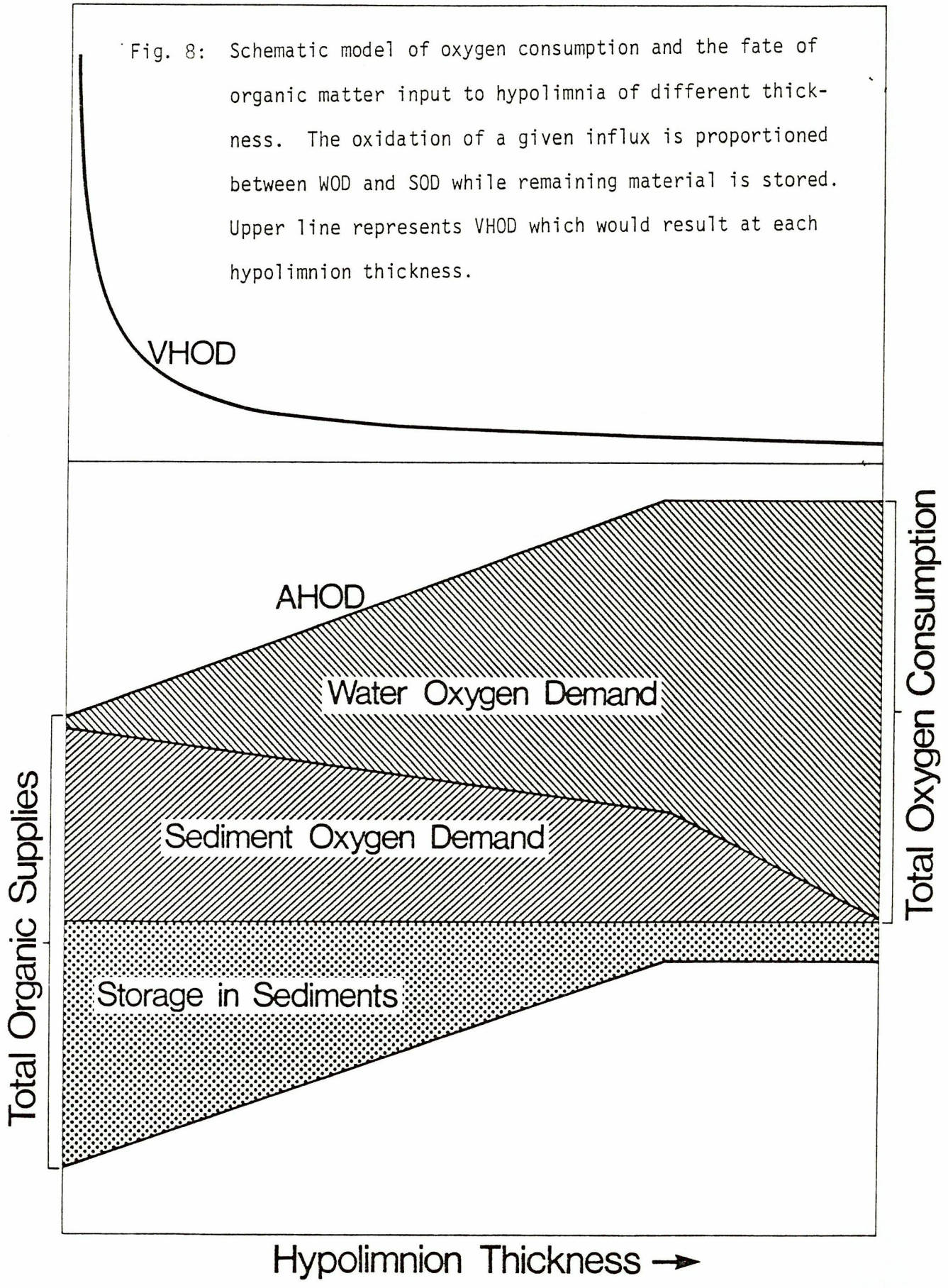


Fig. 8: Schematic model of oxygen consumption and the fate of organic matter input to hypolimnia of different thickness. The oxidation of a given influx is proportioned between WOD and SOD while remaining material is stored. Upper line represents VHOD which would result at each hypolimnion thickness.

warm hypolimnion will have faster depletion rates than a deep cold one.

In shallow hypolimnia the influence of sediment oxygen demand is greater, less is oxidized in water column, but reaches a maximum value when SOD is limited by the diffusion rate of O_2 (Hargrave 1969). At this point storage is more important. These factors are summarized on Fig 8 .

The use of total phosphorus load to predict lake phosphorus concentrations and chlorophyll has been useful in most cases. If however rivers contain large amounts of unavailable P then the impact of a sewage plant will be grossly overestimated. Two examples illustrate this point.

A significant fraction of the P entering Lake Erie is from bluff erosion. All of this is unavailable for plant growth. Most of the input to Kamloops Lake is from the inflowing river. Phosphorus recovery in the sewage plant and a local paper mill seem insignificant. Careful analysis (Daley et al, 1979) showed that most of the river load is apatite P and it either settles out in the lake or passes through unchanged. It is certainly unavailable. This means that much of the bioavailable P was coming from the sewage plant and paper mill.

Our goal must be to monitor not just total P but forms of P that can be used to calculate available P load. In soils only some of the total P is available for plant uptake and extractions, eg. using bicarbonate (Olsen 1954), provide a much better estimate of available P. More recent work (Williams et al, 1967) on the chemical nature of the different forms of phosphorus in soils has shown that apatite is only slowly released to plants but the non apatite inorganic forms of orthophosphate (associated with iron and aluminum) are readily available. Organic P forms are not immediately available for plant growth.

Golterman (1977) had surprising good luck in correlating algal growth and the P extracted by nitrilotriacetic acid (NTA) from sediments. Sagher (1976) found a sodium hydroxide extraction to be a useful way to determine the available P in sediments but the P remaining which was measured after hydrochloric acid extraction was not available.

Williams et al (1980) used the above extractions along with a Scenedesmus algae bioassay and found that the uptake was related to the amount of nonapatite inorganic phosphorus (NAIP). Bluff materials contained over 90% apatite and would not support algal growth. NAIP was correlated with the NaOH and NTA extractable forms and organic P was not used. Uptake rates of NAIP were similar to those for orthophosphate.

About 75% of the NAIP was utilized. This should be considered as only potentially available since culture conditions represent optimum contact between the algae and the source of P. In lakes some of the particulate P may settle out and become buried in sediments.

In view of Williams et als work the EPA method of ignition must be used cautiously. The organic P forms would be measured but are probably unavailable and apatite P would be a least partly extracted and measured but be unavailable.

Lee et al (1980) provide a detailed review and neglecting the strong bias toward other Lee et al publications does provide some valuable insights. The P export was measured for a wide variety of drainage areas. Rarely is the P available to algal growth greater than about 40%. A figure of 35% was considered average. They come up with a useful equation to summarize their findings:

Available P = SRP + 0.2 particulate P

Bioassays are however biased in that they measure forms available over the short term. They should be done but some areas of confusion may result. These are listed below:

1) The presence of a large quantity of refractory P may only be releasing P slowly but since the quantity is large can supply sufficient P to maintain a system for many years.

2) All bioassays fail to consider the potential importance of UV radiation. This may convert some of the "unavailable" organic P forms to orthophosphate.

3) Much of the phosphorus settles out as organic P and is relatively unavailable. As it is slowly decomposed by the benthos the organic P is converted to NAIP which may or may not be available depending on the reducing state of the sediments.

4) It is interesting to speculate on influence of a passage of calcium phosphate type materials through the acidic guts of worms. They process the entire surface zone of sediments several times a year. This may make some unavailable material available.

5) Phosphate esters are not used if orthophosphate is present or if phosphatase activity has not been induced at the time of the test. This would give a false impression that organic esters were unavailable.

In another type of bioassay a series of nutrients are added to lake waters and it is expected that the photosynthesis will show an immediate response to enrichment of the limiting nutrient. Don't use this technique because the opposite happens. The 14 C uptake is depressed as the cells opt for nutrient uptake (Healey 1979, Lean and Pick in press).

The internal loading of P to lakes from sediments is a most contraversal topic. My bias comes from limnocoral experiments and can be summarized as follows. In shallow systems 0-4m where the light does not penetrate to the sediments and where the water is not thermally stratified the phytoplankton can derive most of their nutrient requirements from the sediments. Radioisotope experiments have shown that there is a constant loss of P to the sediments and this is equal to that caught in sediment traps (Lean et al, 1975, Charlton 1975, Lean and Charlton 1976). Balancing this is a continuous return from the sediments keeping the water column in a pseudo- steady state. Additions of N and P can enhance the chlorophyll level and superimposed on this are events which cause a net internal loading. The chemical measurements of phosphorus show massive return of phosphate from the sediments. This triggers the August bloom of phytoplankton in shallow lakes and the input comes from the NAIP form.

Deeper stratified columns of water cause the euphotic zone to be cut off from the nutrient rich hypolimnion and sediments. Here internal loading is of little consequence except in large lakes under storm conditions. It should be remembered that even deep lakes have shallow zones where P uptake and release can proceed. Circulation patterns bring mid-lake plankton to these littoral zones.

I hope I have shown that phosphorus acts as the pacemaker in lakes and nitrogen can have a moderating effect. Reducing phosphorus inputs will have dramatic effects in some lakes, but most of the Great Lakes have a low algal biomass anyhow. In these lakes, improvements due to phosphorus reductions will be most dramatic in the near shore zones and indeed this is where the serious problems exist. On the other hand most of the mid-lake waters, where only a few fishermen or sailers have been, are clear. Now is the time to learn to appreciate the excellent health and beauty of Canadian waters in general, while maintaining proper scientific activities designed to keep them that way.

References

1. Charlton, M.N., 1975: Sedimentation: Measurements in experimental enclosures. - Verh. Internat. Verein. Limnol. 19, 267-272.

2. Charlton, M.N. 1979. Hypolimnetic oxygen depletion in Central Lake Erie: Has there been any change? Inland Waters Directorate Scientific Series No. 110, Ottawa, 1979.

3. Charlton, M.N. 1980. Oxygen depletion in Lake Erie: Has there been any change? Can. J. Fish. Aquat. Sci. 37:000-000.

4. Cornett, R.J. and F.H. Rigler. 1979. Hypolimnion oxygen deficits: their prediction and interpretation. Science 205:580-581.

5. Dillon, P.J. and F.H. Rigler 1974a. The Phosphorus - Chlorophyll relationship in lakes. Limnol. Oceanogr. 19:767-773.

6. Dillon, P.J. and F.H. Rigler 1974b. A test of a simple nutrient budget model predicting the phosphorus concentration in lake water. J. Fish. Res. Bd. Can. 31:1771-1778.

7. Dillon, P.J. and W.B. Kirchner 1975. The effects of geology and land use on the export of phosphorus from watersheds. Water Research 9:135-148.

8. Golterman, H.L. 1977. Sediments as a source of phosphate for algal growth. In H.L. Golterman (ed). Interactions between sediments and fresh water. Junk p 286-293.

9. Hargrave, B.T. 1969. Similarity of oxygen uptake of benthic communities. Limnol. Oceanogr. 14:801-805.

10. Hutchinson, G.E. 1938. On the relation between the oxygen deficit and the productivity and typology of lakes. Int. Rev. Gesamten Hydrobiol. 36:336-355.

11a. Hutchinson, G.E. 1957. A Treatise on Limnology, Vol. 1. J. Wiley and Sons, Inc., New York, 1015pp.

11b. Kirchner, W.B. and P.J. Dillon. 1975. An empirical method of estimating the retention of phosphorus in lakes, Water Resources Res. 11:182-183.

12. Lean, D.R.S. Charlton, M.N. Burnison, B.K. Murphy, T.P. Millard, S.E. & Young, K.R. 1975. Phosphorus: Changes in ecosystem metabolism from reduced loading. - Verh. Internat. Verein. Limnol. 19:249-257.

13. Lean, D.R.S. & Charlton, M.N. 1976. A study of phosphorus kinetics

in a lake ecosystem. In. Nriagu, J.O. (ed) Environmental Biogeochemistry. Vol. 1:283 - 294. Ann Arbor Science, Ann Arbor, Mich.

14. Lean, D.R.S. and F.R. Pick (in press). Photosynthetic response by lake plankton to nutrient enrichment: A re-evaluation of the ^{14}C radiobioassy.

15. Leppard G.G., A. Massalski, and D.R.S. Lean. 1977. Electron-opaque microscopic fibrils in lakes: their demonstration, their biological derivation and their potential significance in the redistribution of cations. Protoplasma 92:289-309.

16. Lee, G.F. R.A. Jones, and W. Rast. 1979. Availability of phosphorus to phytoplankton and its implications for phosphorus management strategies. Proc. Cont. Management Strategies for the Great Lakes. April 1979. International Joint Commission Pub.

17. Massalski, A., and G.G. Leppard. 1979. Survey of some Canadian lakes for the presence of ultrastructurally discrete particles in the colloidal size range. J. Fish. Res. Board Can. 36: 906-921.

18. Massalski, A., and G.G. Leppard. 1979. Morphological examination of fibrillar colloids associated with algae and bacteria in lakes. J. Fish. Res. Board Can. 36:922-938.

19. Olsen, S.R., C.V. Cole, F.S. Watanabe, and L.A. Dean. 1954. Estimation of available phosphorus in soils by extraction with sodium bicarbonate. U.S. Dep. Agric. Circ. 939

20. Rawson, D.S. 1955. Morphometry as a dominant factor in the productivity of large lakes. Verh. Internat. Verein. Limnol. 12:164-175.

21. Verein. Limnol. 19:197-210

22. Ryder, R.A. 1964. Chemical characteristics of Ontario Lakes with reference to a method for estimating food production. Section report. (Fisheries) No. 48, Dept. of Lands and Forests, Ontario 75p.

23. Sagher, A. 1976. Availability of soil run off phosphorus to algae. Ph.D. Thesis. U. Wisconsin - Madison. 192p.

24. Sakamoto, M. 1966. Primary production by phytoplankton community in some Japanese lakes and its dependence on lake depth. Arch. Hydrobiol. 62:1-28.

25. Thienemann, A. 1928. Der Sauerstoff im Eutrophen und Oligotrophen see. Die Binnengewasser, 4. Stuttgart, Schweizerbartsche Verlagsbuchhandlung, 175 pp.

26. Williams, J.D.H., J.K. Syers, and T. W. Walker, 1967. Fraction of soil inorganic phosphate by a modification of Chang and Jacksons procedures. Soil. Sci. Am. Proc. 31: 736-739.

27. Williams, J.D.H., H. Shear, and R.L. Thomas. 1980. Availability to Scenedesmus guadricauda of different forms of phosphorus is sedimentary materials from the Great Lakes. Limnol. Oceanogr. 25:1 - 11.

28. Vollenweider, R.A. 1968. The scientific basis of lake and stream eutrophication with particular reference to phosphorus and nitrogen as eutrophication factors. Tech. Rep. OECD, Paris, DAS/C81/68, 27:1-182.

NUTRIENT SOURCES

by

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INTRODUCTION

Cultural eutrophication is caused by excessive loads of aquatic plant nutrients to natural waters which can produce nuisance growths of algae and higher aquatic plants. While some waters are naturally eutrophic, an increased nutrient load has often resulted from human activity in the drainage area surrounding the water body.

Because compounds of nitrogen and phosphorus are frequently the primary nutrients that stimulate undesirable algal growth, the object of this paper is to identify the major sources of these nutrients in the aquatic environment. This approach is taken with the realization that in some circumstances nutrients other than nitrogen and phosphorus may actually be more important in controlling receiving water quality. Although point source control of nitrogen and phosphorus is the subject of this seminar, some consideration will be given to the magnitude of diffuse nutrient inputs. Hopefully, this will more clearly illustrate the importance of nutrient control from point discharges.

PHOSPHORUS

In the Great Lakes, phosphorus has been found to be the nutrient which most often limits aquatic plant growth. A comparison of several water quality parameters (Figure 1) indicates that a deterioration of the trophic status of a lake can often accompany increased phosphorus loadings. Phosphorus is also the nutrient most easily controlled by alterations in wastewater treatment practices, detergent reformulation, and in the collection and disposal of runoff from urban and agricultural land.

Diffuse Sources

Until recently, the Great Lakes were considered to be an inexhaustible source of high quality fresh water. However, it is now recognized that the assimilative capacity of these lakes is finite and that the advanced trophic status of Lakes Erie and Ontario is the result of an historical disregard for phosphorus loadings. Consequently, the International Joint Commission established an intensive survey of pollutant sources by the International Reference Group on Great Lakes Pollution from Land Use Activities (PLUARG, 1978). The final report of the Reference Group contains considerable information regarding the contributions of various land use types to the total phosphorus loadings in the Great Lakes Basin. Although the study pertains directly to the Great Lakes watershed, the expression of phosphorus loadings by land use types may permit application of the same relationships to other jurisdictions.

A summary of the 1976 total phosphorus loads for a relatively oligotrophic lake (Huron) and a eutrophic lake (Erie) is presented in Table 1. A comparison of the point and non-point sources of the total phosphorus load can be made upon examination of this table. Diffuse tributary inputs comprise a large proportion of the total loads to the lakes, accounting for 50% in Lake Huron and 48% in Lake Erie. Inputs from sewage treatment plants are almost insignificant for Lake Huron, but in the Lake Erie case contribute 32% of the total.

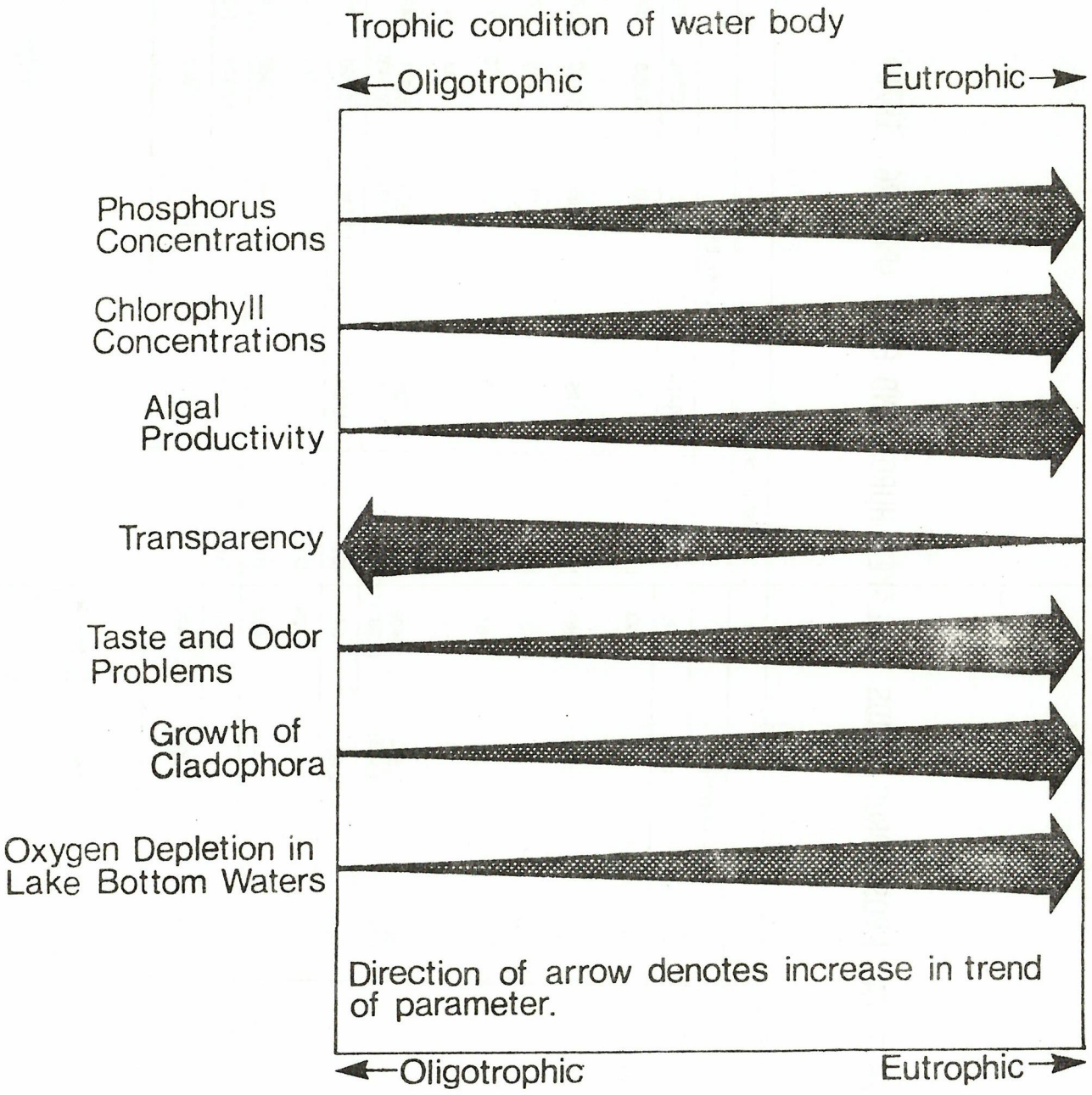


FIGURE 1: EUTROPHICATION RELATIONSHIPS IN LAKES (PLUARG, 1978)

TABLE 1: SUMMARY OF 1976 TOTAL PHOSPHORUS LOADS TO LAKES HURON AND ERIE (PLUARG, 1978)

	metric tons/yr							
	LAKE HURON				LAKE ERIE			
SOURCE	CANADA	U.S.	TOTAL	[PERCENT]	CANADA	U.S.	TOTAL	[PERCENT]
Direct Municipal Sewage Treatments Plants	107	16	123	[3]	70	5,588	5,658	[32]
Tributary Municipal Sewage Treatment Plants	83	309	392	[8]	185	985	1,170	[7]
Direct Industrial	0	31	31	[<1]	164	111	275	[2]
Tributary Industrial	0	81	81	[2]	0	72	72	[<1]
Urban Nonpoint Direct	16		16	[<1]	44		44	[<1]
Tributary Diffuse (Tributary Total)	864 (947)	1,564 (1,954)	2,428 (2,901)	[50]	1,726 (1,911)	6,675 (7,732)	8,401 (9,643)	[48]
Sub-Total	1,070	2,001	3,071	[63]	2,189	13,431	15,620	[89]
Atmospheric	—	—	1,129	[23]	—	—	774	[4]
Load From Upstream Lake	—	—	657	[14]	—	—	1,080	[6]
Total			4,857	[100]			17,474	[100]
Shoreline Erosion (Not Included in Total)	131	295	426		5,912	1,024	6,936	

Phosphorus from shoreline erosion was not included in total lake phosphorus loading estimates. PLUARG studies indicated that shoreline phosphorus was almost exclusively apatite phosphorus. Under the pH conditions normally existing in lakes, none of this phosphorus would be available for biological growth.

Comparisons and ranges of unit area phosphorus loads by dominant land use are given in Table 2. Unit area loads are obtained by dividing total contributions from a given land area by the size of the land area. Unit area loads are useful for the comparison of non-point contributions between different land uses. The data of Table 2 are based on two years of intensive monitoring for several types of pollutants.

The results indicate that unit area loads of phosphorus from agricultural and urban land uses are approximately equal with both being one to two orders of magnitude greater than those of forested and/or idle land. Forested and idle land unit area loads are considered to be at or near background levels. By far the largest unit area contribution is associated with developing urban land. The developing area category, which includes soil-exposing construction sites, is of major concern, not only in the Great Lakes Basin, but probably also in any location where the development of agricultural land for urban use is significant.

To gauge the contribution to overall phosphorus loading from diffuse sources, a more comprehensive breakdown of unit area loadings can be utilized. In Table 2, unit area loads were categorized only by land use type. Tables 3 and 4 take the process one step further by incorporating the type of soil for agricultural use and the degree of industrialization for urban areas. Table 3 indicates that although unit area loads for a given land use may vary by a factor of ten or more, knowledge of soil type permits a more refined estimate to be made.

Phosphorus unit area loads in urban areas rise with increasing industrialization (Table 4) and are much higher than those from rural areas. The load from urbanizing land is particularly high.

TABLE 2: SUMMARY OF RANGES OF UNIT AREA LOADS OF SELECTED MATERIALS BY LAND USE (PLUARG, 1978)

Land Uses	ANNUAL UNIT AREA LOADS (kg/ha/yr)							
	Suspended Solids	Total Phosphorus	Filtered Reactive Phosphorus	Total Nitrogen	Lead	Copper	Zinc	Chloride
I. **RURAL:**								
General Agriculture	3 – 5,600	0.1 – 9.1	0.01 – 0.6	0.6 – 42	0.002 – 0.08	0.002 – 0.09	0.005 – 0.3	10 – 120
Cropland	20 – 5,100	0.2 – 4.6	0.05 – 0.4	4.3 – 31	0.005 – 0.006	0.014 – 0.064	0.026 – 0.083	10 – 50
Improved Pasture	30 – 80	0.1 – 0.5	0.02 – 0.2	3.2 – 14	0.004 – 0.015	0.021 – 0.038	0.019 – 0.172	—
Forest/Wooded	1 – 820	0.02 – 0.67	0.01 – 0.10	1 – 6.3	0.01 – 0.03	0.02 – 0.03	0.01 – 0.03	2 – 20
Idle/Perennial	7 – 820	0.02 – 0.67	0.01 – 0.07	0.5 – 6.0	0.01 – 0.03	0.02 – 0.03	0.01 – 0.03	20 – 35
Sewage Sludge	—	.02	0.01	11	0.01	0.005	0.2	10
Wastewater Spray Irrigation	--	0.2 – 1.4	0.1 – 1.3	2.2 – 370	—	—	—	40 – 160
II. **URBAN:**								
General Urban	210 – 1750	0.3 – 2.1	0.05 – 0.3	6.2 – 10	0.14 – 0.5	0.05 – 0.13	0.3 – 0.6	130 – 380
Residential	620 – 2,300	0.4 – 1.3	0.2	5 – 7.3	0.06	0.03	0.02	1,050
Commercial	50 – 830	0.1 – 0.9	0.02 – 0.08	1.9 – 11	0.17 – 1.10	0.07 – 0.13	0.25 – 0.43	10 – 150
Industrial	400 – 1,700	0.9 – 4.1	0.3	1.9 – 14	2.2 – 7.0	0.29 – 1.3	3.5 – 12.0	—
Developing Urban	27,500	23	0.1	63.0	—	—	—	75 – 160

TABLE 3: PREDICTED TOTAL PHOSPHORUS UNIT AREA LOADS FOR RURAL LAND, FORESTED LANDS AND WETLANDS (PLUARG, 1978)

	kg/ha/yr					
	Type of Soil					
Land Use Intensity	Sand	Coarse Loam	Medium Loam	Fine Loam	Clay	Organic
Rural						
Row Cropping (> 50 percent row crops)	0.25	0.65	0.85	1.05	1.25[b]	—
Mixed Farming (25 – 50 percent row crops)	0.10	0.20	0.30	0.55	0.85	—
Forage (< 25 percent row crops)	0.05	0.05	0.10	0.40	0.60	—
Grassland	0.05	0.05	0.10	0.15	0.25	—
Forest	0.05	—	—	—	0.10[c]	—
Wetlands						
Natural areas	—	—	—	—	—	0
Cultivated Organic Soils	—	—	—	—	—	2.20

TABLE 4: PREDICTED TOTAL PHOSPHORUS URBAN UNIT AREA LOADS (PLUARG, 1978)

	kg/ha/yr		
	Degree of Industrialization		
Urban	Low	Medium	High
Combined Sewered Areas	9	10	11
Separate Sewered Areas	1.25	2.5	3.0
Unsewered Areas	1.25	—	—
Small Urban Areas (Sewer System Not Differentiated)	2.5	2.5	2.5
Urbanizing Land	25	25	25

Municipal Point Sources

In Ontario, the major point sources of phosphorus are municipal wastewater treatment plants. The approach taken by PLUARG to quantify this contribution was to estimate an average per capita phosphorus production and then apply a reduction factor corresponding to the phosphorus removal efficiency achieved in water pollution control plants.

During the late 1960's, when the phosphorus content of household detergents was unregulated, the per capita discharge of phosphorus in municipal wastewater was relatively high. Zanoni and Rutkowski (1972) reported that with high phosphate detergents, the per capita phosphorus loadings were 3.3 to 4.8 kg a^{-1}. A later study carried out at the Canadian Forces Base, Uplands, by Stepko and Shannon (1974) found that per capita production was 1.9 kg a^{-1}. Weighted averages from 94 Ontario water pollution control plants were used to construct Figure 2 (Appleby, 1977). Before 1970, detergent phosphorus content averaged 11% as P. The Canadian Government limited phosphorus content of laundry detergents to 8.7% in 1970 and 2.2% on January 1, 1973. The resulting effect on sewage phosphorus concentrations is shown in Figure 2. The dip in 1974 and 1975 is associated with a shortage of phosphorus and a depressed economy during this period. It is of interest to note that detergent reformulation coupled with the introduction of chemical phosphorus removal in Ontario has reduced per capita loadings in control plant effluents from 1.1 kg a^{-1} in 1969 to less than 0.2 kg a^{-1} in 1978.

Detergent reformulation was also found to have a pronounced effect on the diurnal fluctuation of total phosphorus in raw sewage. Average diurnal variations in influent wastewater phosphorus before and after formulation are shown in Figure 3. While the per capita contribution was reduced by 56%, the amplitude of the diurnal fluctuation was also decreased. The reduced variability in wastewater phosphorus concentration makes the addition of phosphorus removal chemicals in proportion to changing phosphorus loadings a much simpler task.

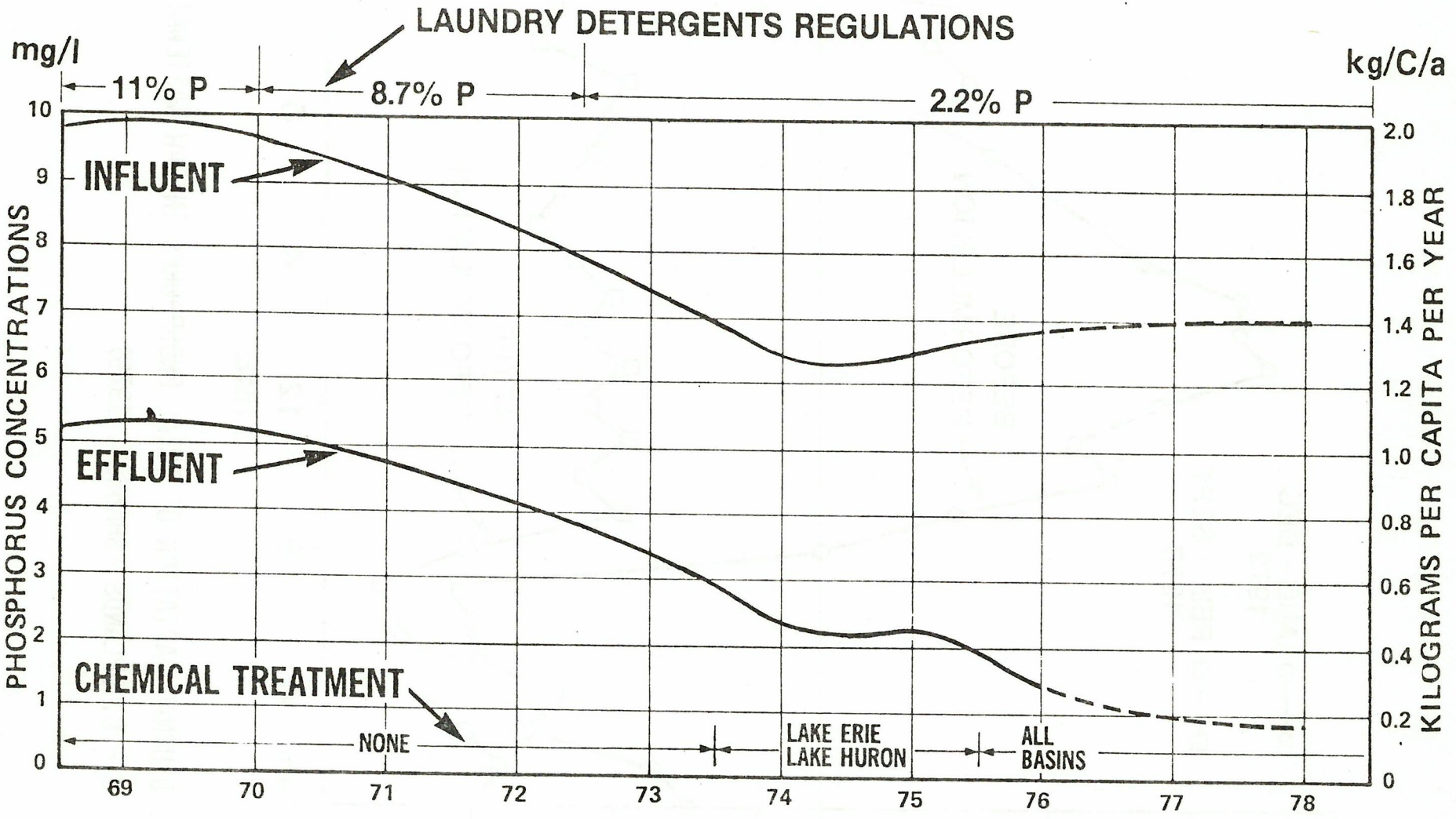


FIGURE 2: PHOSPHORUS CONCENTRATIONS IN ONTARIO WASTEWATER (APPLEBY, 1977)

dip in 74 due to poor economy.

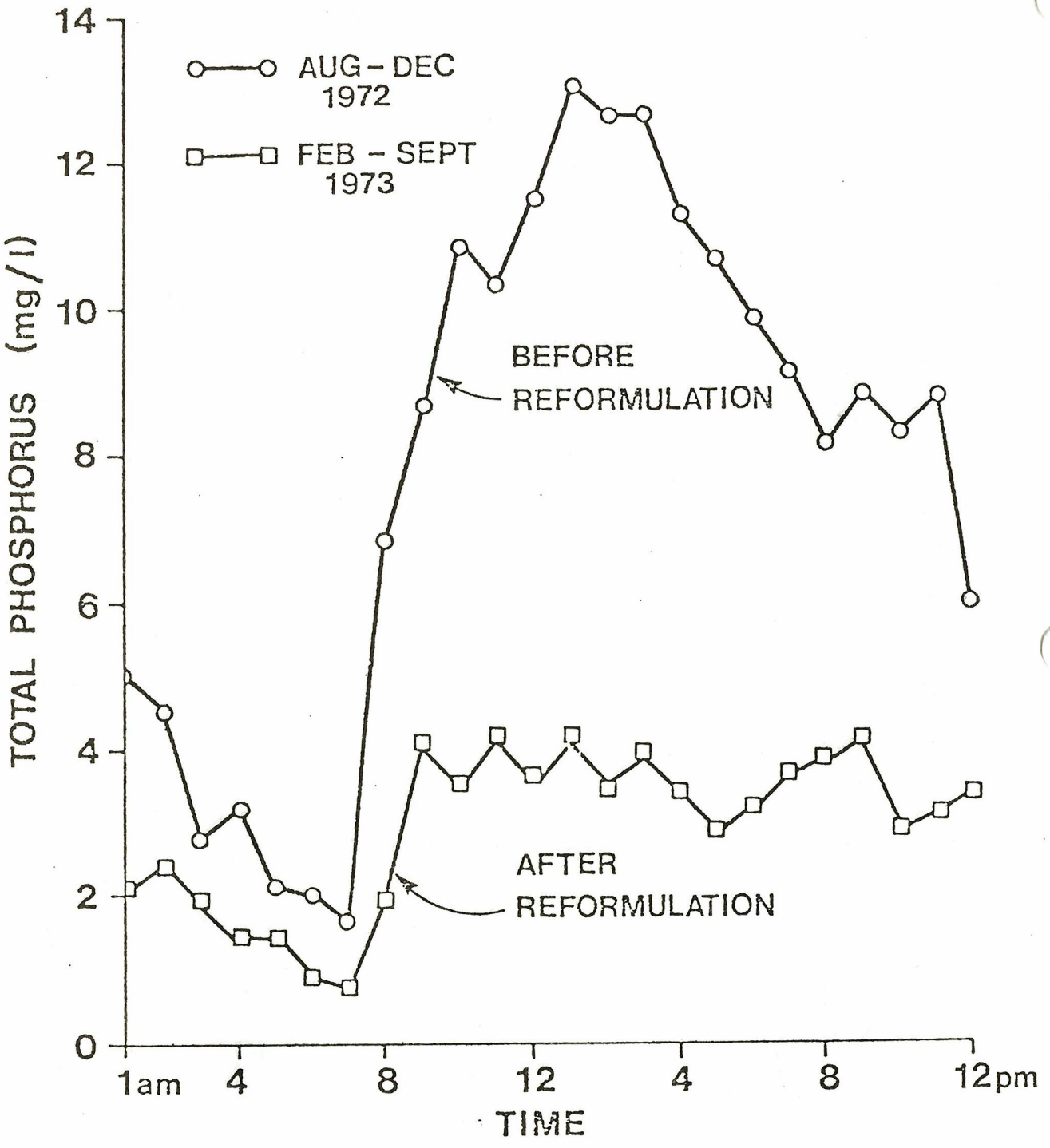


FIGURE 3: DIURNAL VARIATION OF TOTAL PHOSPHORUS IN RAW WASTEWATER AT C.F.B. UPLANDS (SHANNON, 1979)

The estimated sources of phosphorus in municipal sewage are indicated in Figure 4. Cleaning compounds, exclusive of laundry detergents, include household and industrial dishwashing compounds, metal cleaning compounds, water treatment compounds and all other industrial and institutional cleaning compounds. The phosphorus loadings from all cleaning compounds, including laundry detergent, has decreased from 47% of the total in 1969 to 29% in 1976, primarily as a result of detergent reformulation (Table 5).

TABLE 5. PHOSPHORUS CONTRIBUTION FROM CLEANING COMPOUNDS IN CANADA (APPLEBY, 1977)

	1969	1971	1973	1975
Canadian population (10^6)	20.7	21.4	22.1	22.8
Canadian phosphorus sales to laundry detergents (Gigagrams)	11.8	7.6	2.6	2.8
grams P/capita/annum	570	350	120	120
Phosphorus sales to the rest of the cleaning compounds (g/C/a)	360	370	330	250
TOTAL - g/C/a	930	720	450	370
Estimated concentration in raw sewage - mg/L	4.7	3.6	2.3	1.9

The contribution from human waste is reported to be between 500 and 600 grams of phosphorus per capita per year. Of this amount, about 400 grams is in urine and the remaining portion in faeces (Table 6).

The miscellaneous category, including kitchen wastes, is defined as the difference between the known total inputs and estimated contributions due to human waste and Canadian phosphorus production. Some of the components of this category are imported organic and inorganic phosphates for non food-grade use such as anti-freeze, toothpaste, fire retardants, pesticides and oil additives.

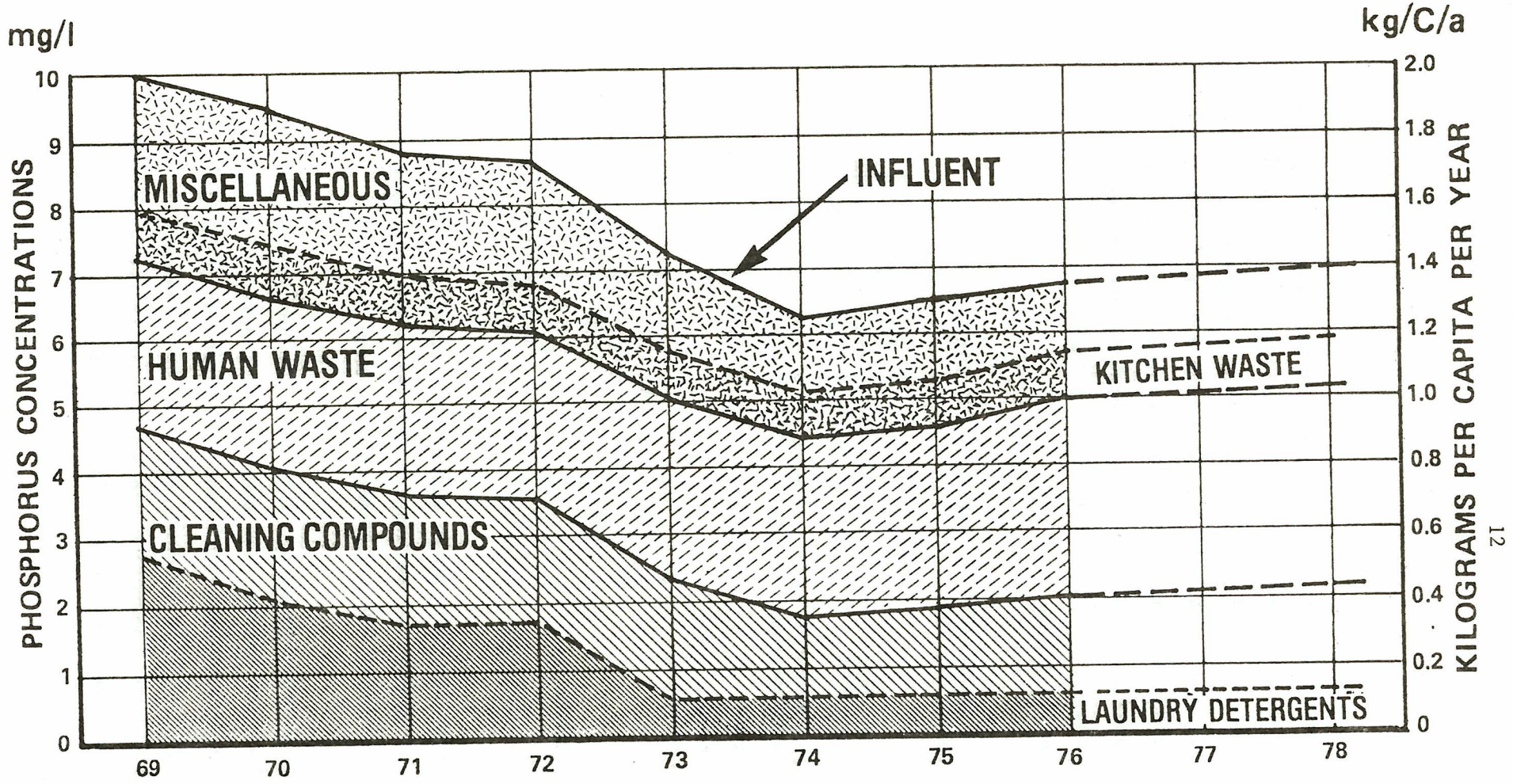


FIGURE 4: SOURCE OF PHOSPHORUS IN RAW SEWAGE IN ONTARIO (APPLEBY, 1977)

TABLE 6. PERCENT CONTRIBUTIONS TO RAW SEWAGE PHOSPHORUS BY HUMAN WASTE AND MISCELLANEOUS SOURCES (APPLEBY, 1977)

	1969	1971	1973	1975	1976
Human wastes	25	29	38	42	43
Cleaning compounds:					
Laundry detergents	28	20	8	9	9
Others	19	22	24	20	20
Sub-total	47	42	32	29	29
Miscellaneous (including kitchen waste at 10%)	28	29	30	29	28
TOTAL	100	100	100	100	100

Only a portion of the total phosphorus in municipal sewage is actually available for biological growth. The biological availability of phosphorus from diffuse sources has attracted considerable interest (Lee et al, 1979), but little is known of phosphorus associated with sewage. The results shown in Figure 5 are from a study which is in progress at the Wastewater Technology Centre, Burlington, Ontario. Phorphorus levels in four process streams of a biological treatment plant utilizing ferric chloride addition for phosphorus removal were analyzed and divided into total, particulate and soluble portions. Each portion was then further sub-divided into different chemically-reacting fractions. Of the total phosphorus in each stream, the fractions generally considered to be potentially biologically available (Williams et al, 1976) are the soluble reactive phosphorus (RP), nonapatite inorganic phosphorus (NAIP) and some portion of the organic phosphorus (OP).

For the case shown in Figure 5(a), approximately 67% of the influent total phosphorus of 6.2 mg L^{-1} was available. After primary treatment (Figure 5(b)), phosphorus concentrations were considerably higher due to the wastage of secondary sludge and digester supernatant to the primary clarifiers. Following bio-chemical treatment, the phosphorus level in the plant effluent was about 1.0 mg L^{-1} and 75% of this amount was available. The increase in the available fraction across the plant indicated that removal efficiency was slightly greater for the non-available than for the available phosphorus.

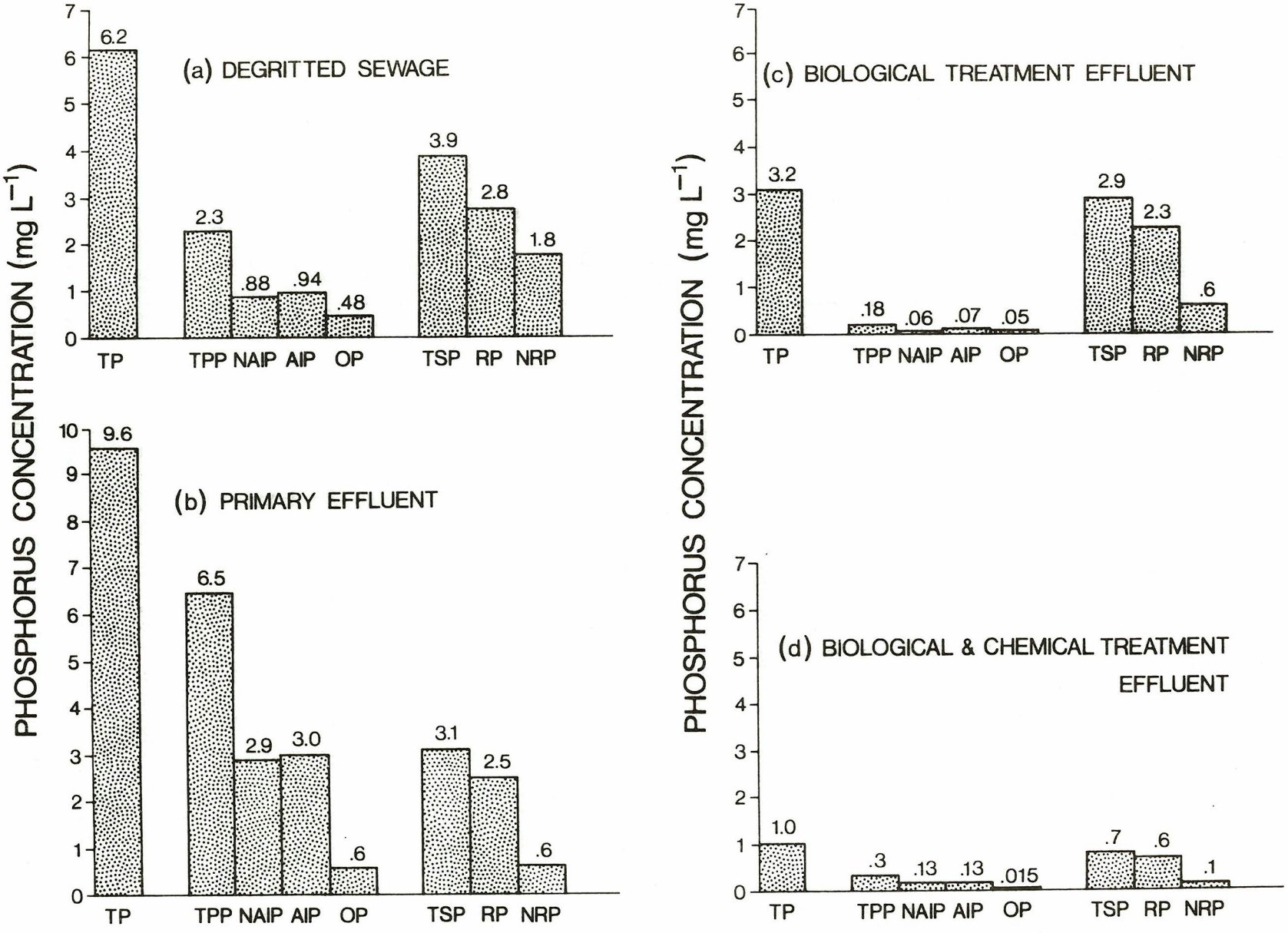


FIGURE 5: PHOSPHORUS FRACTIONS IN PROCESS STREAMS OF BURLINGTON SKYWAY WPCP

(TP= total P; TPP= total particulate P; NAIP= nonapatite inorganic P; AIP= apatite inorganic P; OP= organic P; TSP= total soluble P; RP= reactive P; NRP= non reactive P)

NITROGEN

In addition to its role as an algal nutrient in receiving waters, nitrogen is of concern for several other reasons (U.S. EPA, 1975) which include:

(a) ammonia at low concentrations is toxic to fish;
(b) ammonia is corrosive to material such as copper;
(c) ammonia increases breakpoint chlorination requirements for water pollution control plant effluents;
(d) reduced nitrogen exerts a large oxygen demand on receiving water; and
(e) nitrates in drinking water have been implicated in infant methemoglobinemia.

Nitrogenous compounds can enter the aquatic environment from both natural and human related sources. The major natural sources are precipitation, dustfall, non-urban runoff and biological nitrogen fixation. Several of these categories are stongly affected by man's activities. For example, fossil fuel combustion and high nitrogen fertilizers can increase rainfall and non-urban runoff concentrations, respectively.

Diffuse Sources

Unit area loadings of total nitrogen for watersheds of various land uses were given in Table 2 (PLUARG, 1978). In contrast to phosphorus, the average nitrogen loadings for every land use type are generally at least an order of magnitude greater. The difference in N/P ratio for the agricultural area is not surprising in view of the fact that agriculture is the greatest consumer, not only of commercially produced nitrogen, but also of sewage-derived nitrogen used for fertilizer.

Runoff from farm feedlots is an important rural nitrogen source, especially during and following a rainfall. The greatest pollutant concentrations occur during periods of warm weather and low rainfall intensity. Ammonia nitrogen from urea hydrolysis is the major feedlot nitrogen component and may reach concentrations of 150 to 300 mg L^{-1} in runoff. Organic nitrogen concentrations of up to 600 mg L^{-1} have also been reported (U.S. EPA, 1975).

Urban runoff is a highly variable nitrogen source. As with feedlot operations, the highest nutrient concentrations are usually associated with rainfall occurring after a lengthy dry period. Under these conditions, runoff from a residential and light commercial area may result in total nitrogen concentrations of 0.3 to 7.5 mg L^{-1} (AWWA, 1967).

Point Sources

The importance of point sources of nitrogen to the receiving water near a highly urban area is illustrated in Table 7. In the San Francisco Bay Basin, 79% of the nitrogen emissions are associated with point sources, both municipal and industrial. Therefore, any strategy for the management of nitrogen pollution must include nitrogen control for point source discharges.

TABLE 7. ESTIMATED NITROGEN LOADINGS FOR THE SAN FRANCISCO BAY BASIN (U.S. EPA, 1975)

Source	Nitrogen Load kg a^{-1} ($x10^{-3}$)	Percent of Total
Municipal Wastewater (untreated)	26 000	49
Industrial Wastewater (untreated)	16 000	30
Vessel Wastes (untreated)	60	0.1
Dustfall on Bay	590	1.1
Rainfall on Bay	390	0.8
Urban Runoff	1 400	2.7
Non-urban Runoff	1 900	3.6
Nitrogen from Agricultural Fertilizer	900	1.7
Feedlots and Dairies	6 000	11
TOTAL:	53 000	100

The nitrogen content of industrial wastes may be highly variable and is usually dependent on the nature, location and size of the industry. Table 8 indicates the magnitude of the major sources of waste nitrogen from industrial sources in Canada. Fertilizer manufacturing utilizes large quantities of nitrogen as a raw material and

generally produces concentrated nitrogenous wastes. In Canada, 70% of the waste nitrogen discharged in this manner is emitted in the western provinces where fertilizer manufacturing is prominent.

TABLE 8. SOME INDUSTRIAL SOURCES OF WASTE NITROGEN IN CANADA (BRIDLE, 1979)

Industry	Discharge (kg d^{-1})		
	NH_3-N	NO_3-N	TKN
Fertilizer	9 300	7 800	17 100
Iron & Steel	-	-	19 000
Org. Chemicals	-	-	3 000

The iron and steel industry produces more than 19 000 kg d^{-1} of waste nitrogen. Most of these emissions are associated with coking operations in which ammonia and organic nitrogen are released during the distillation of coal. The nitrogen content of coal is about 2% w/w, but during coking only a small portion of this amount is lost - the rest remains with the coke (AWWA, 1967).

Significant quantities of nitrogen waste are also associated with the petrochemical industry whose feedstock, oil, contains roughly the same quantity of nitrogen as coal. Petroleum refineries and the chemical industry often generate large volumes of wastes that are high in nitrogen (AWWA, 1967; Bridle et al, 1979). The ammonia nitrogen content of condensate waters from petroleum refining is frequently hundreds to several thousands of milligrams per litre.

Other industrial sources include meat packing and processing plants, milk processing plants, ice plants and industries using ammonia for scouring and cleaning (U.S. EPA, 1975).

The primary sources of nitrogen in domestic wastes are faeces, urine and waste food. The substitution of sodium nitrilo-triacetate for phosphate as the prime detergent builder in Canada has introduced another nitrogen source. Per capita nitrogen loading lies

in the range of 3.6 to 5.4 kg a^{-1} and is roughly equally divided between ammonia and organic nitrogen (AWWA, 1967). About 50% of the organic nitrogen component can be hydrolyzed biologically to ammonia either during transport of the wastes in the sewerage system or during treatment. Much of the organic nitrogen is contained in suspended solids which can be removed by a solids settling operation.

Nitrogen concentrations in municipal wastes generally range from 15 to 50 mg L^{-1}. As was shown for phosphorus, nitrogen concentrations are also subject to a diurnal variation (Figure 6). Randtke and McCarty (1977) have indicated that peaks in ammonia concentration are usually associated with high flow rates and when taken together result in even greater peaks in ammonia mass loadings to a biological treatment plant.

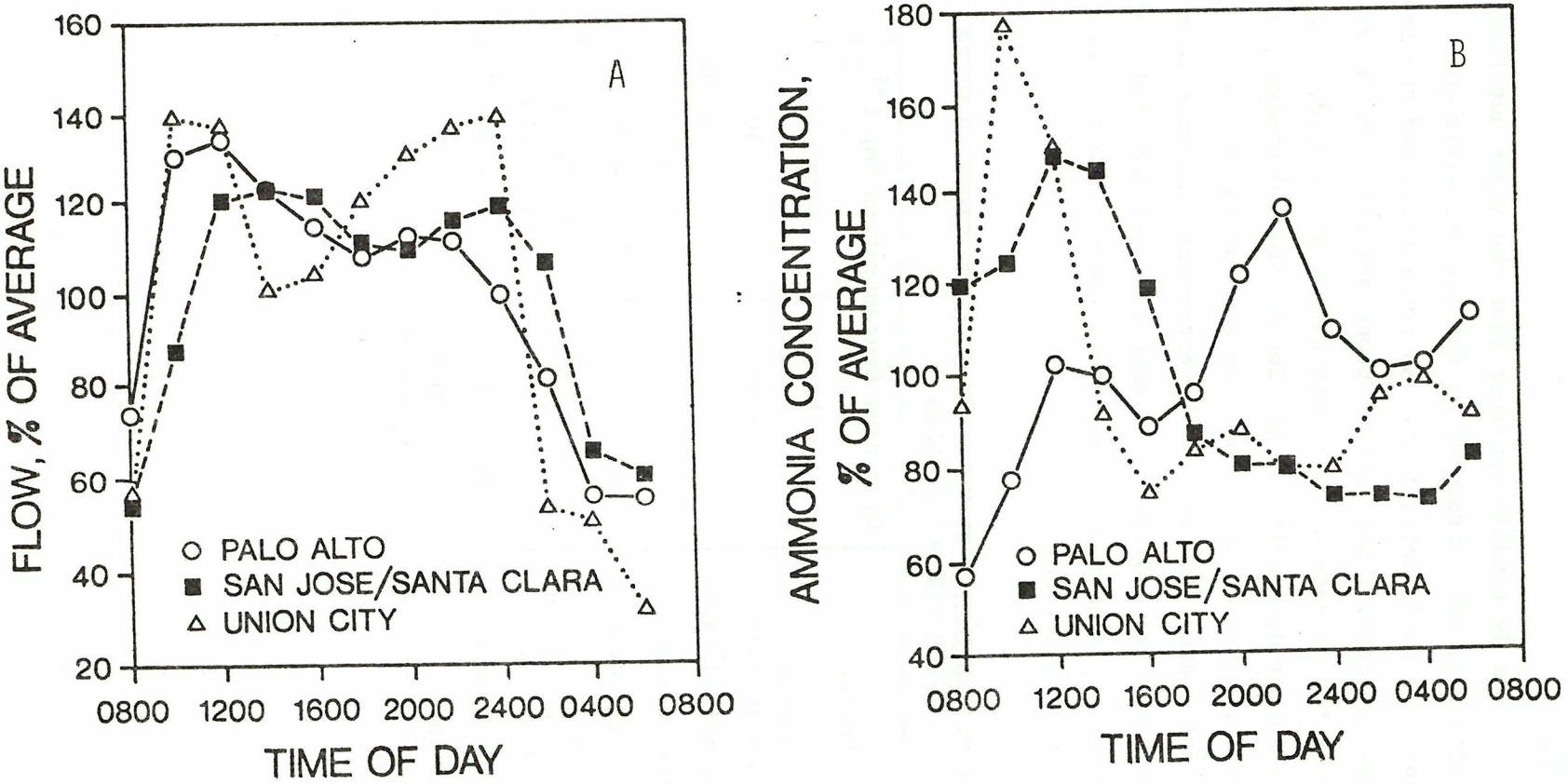


FIGURE 6: NORMALIZED DIURNAL VARIATION IN FLOW (A) AND AMMONIA CONCENTRATION (B) AT THREE WATER POLLUTION CONTROL PLANTS (RANDTKE AND MCCARTY, 1977)

SUMMARY

It is readily apparent that the major nutrients, phosphorus and nitrogen, are introduced to the aquatic environment principally as a result of man's activities. As populations and the level of technology increase, these contributions are likely to be even greater. Because the sources of nutrients are so widespread, complete control would appear difficult, if not impossible. Obviously, the best immediate approach is to focus on controlling the most concentrated nutrient sources (Table 9). Unfortunately, as experience with the Great Lakes has indicated, in some circumstances additional initiatives may be required to completely prevent the problems associated with eutrophication.

TABLE 9. ESTIMATES OF NUTRIENT CONCENTRATIONS IN WATER FROM VARIOUS SOURCES

Source	Discharge Concentration (mg L^{-1})	
	Phosphorus	Nitrogen
Domestic Waste	3 - 10	15 - 50
Industrial Waste	-	0 - 10 000
Rural Runoff:		
Agricultural	0.05 - 1.1	1 - 70
Idle	0.04 - 0.2	0.1 - 0.5
Feedlot Waste	-	16 - 140
Urban Runoff	0.1 - 1.5	1 - 10
Rainfall	0.01 - 0.01	0.1 - 2.0

REFERENCES

Appleby, D.J., "The Impact of Phosphorus Control Activities. The Experience in Ontario", ERCO Industries Ltd., Islington, Ontario, October, 1977.

AWWA Task Group Report, "Sources of Nitrogen and Phosphorus in Water Supplies", Jour. Am. Water Works Assoc., 59:344, 1967.

Bridle, T.R., Personal Communication, 1979.

Bridle, T.R., D.C. Climenhage and A. Stelzig, "Operation of a Full Scale Nitrification-Denitrification Industrial Waste Treatment Plant", Jour. Water Pollut. Control Fed., 51:127, 1979.

Lee, G.F., R.A. Jones and W. Rast, "Availability of Phosphorus to Phytoplankton and It's Implications for Phosphorus Management Strategies", Presented at 11th Cornell University Conf. on Phosphorus Management Strategies for the Great Lakes, Rochester, New York, April, 1979.

PLUARG, "Environmental Management Strategy for the Great Lakes System", Final Report to the Int'l Joint Commission, from the Int'L Reference Group on Great Lakes Pollution from Land Use Activities, July, 1978.

Randtke, S.J. and P.L. McCarty, "Variation in Nitrogen and Organics in Wastewaters", Jour. Env. Eng'g. Div. ASCE, 103:539, 1977.

Shannon, E.E., "Water Quality and Wastewater Treatment Considerations for Detergent Substitutes", Presented at 11th Cornell University Conf. on Phosphorus Management Strategies for the Great Lakes, Rochester, New York, April, 1979.

Stepko, W.D. and E.E. Shannon, "Phosphorus Removal Demonstration Study Using Ferric Chloride and Alum at CFB Uplands", Environmental Protection Service Report No. EPS 4-WP-74-5, Environment Canada, Ottawa, Ontario, 1974.

U.S. EPA, "Process Design Manual for Nitrogen Control",Office of Technology Transfer, Washington, D.C., 1975.

Williams, J.D.H., J.M. Jaquet and R.L. Thomas, "Forms of Phosphorus in the Surficial Sediments of Lake Erie", Jour. Fish. Res. Board Can., 33:413, 1976.

Zanoni, A.E. and R.J. Rutkowski, "Per Capita Loadings of Domestic Wastewater", Jour. Water Pollut. Control Fed., 44:1756, 1972.

HISTORY OF NUTRIENT CONTROL IN ONTARIO

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HISTORY OF NUTRIENT CONTROL IN ONTARIO

1.0 Concerns

All lakes and rivers experience a life-cycle in which characteristic species and life forms are associated with the environmental conditions encountered in each phase of the natural aging process. Eutrophication is the term commonly used, and is the process whereby a body of water is enriched with aquatic plant nutrients to concentrations which result in nuisance levels of biological growth. This biodynamic life-cycle includes primary producers, the algae and rooted aquatics, predators such as zooplankton, fish and scavengers, as well as bacteria, snails and worms. Within limits, the mass of individual organism specie within an ecosystem is directly proportional to the nutrient concentration. Significant shifts in biological response may occur with increasing nutrient loads.

Young , or oligotrophic lakes are poorly supplied with nutrients and are typically low in biological productivity. As a result, the waters are clear, and deeper waters are well supplied with oxygen throughout the year. Such lakes support little fish life.

As the supply of nutrients is progressively increased, the oligotrophic lake goes through a mesotrophic stage. Biological productivity is increased, first through levels which are optimal for the propogation of desirable game fish until eventually conditions are reached which result in a level of productivity to a point where the lake no longer provides adequate habitat for desired species.

Eutrophic lakes have high biological production, the waters are turbid because of dense phytoplankton growth, and the deeper waters, during periods of restricted circulation may become deficient or even devoid of oxygen due to the decomposition of the organic matter produced. Such conditions lead to the disappearance of desirable aquatic life dependent upon the dissolved oxygen, and to the abundance of pollution tolerating forms.

The most visible and troublesome evidence of the biological responses of the lake to excessive nutrient input is the algae blooms. These blooms can clog water intakes, cause turbidity, obnoxious taste and odour, increase backwashing operation frequency and interfere with coagulation and sedimentation processes at the water plants. They restrict the use of pleasure craft, reduce the open water area for sport fishing and result in odourous and unsightly scum along shore lines.

Natural eutrophication is generally, a slow process requiring hundreds, or even thousands of years for objectionable water characteristics to develop. Cultural eutrophication, caused by man's various activities proceeds much more rapidly. It follows, that the control of accelerated biological overproduction becomes a matter of eliminating or controlling the nutrients (or the critical nutrient) essential for the undesirable propagation and growth.

1.1 Concerns In Ontario

Concern for the quality of the Lower Great Lakes was highlighted with the publication of an International Joint Commission (IJC) report, based on a 6-year study, which described the conditions of Lakes Erie and Ontario as follows (1):

"On the basis of a collective evaluation of the foregoing categories (physical, chemical and biological parameters), it is obvious that the western basin of Lake Erie is eutrophic; the central basin moderately eutrophic, and the eastern basin mesotrophic with some oligotrophic characteristics. Lake Ontario is presently in a stage between oligotrophy and mesotrophy."

Further evidence of the deteriorated condition of Lake Erie was provided by the Joint Canada-United States "Project Hypolimnion" (2). This intensive study of the hypolimnion of the central basin of Lake Erie, in 1970, showed that the area of complete oxygen depletion by the end of the summer reached an extent of 6000 sq km or 40% of the whole central basin. The major cause of the oxygen depletion was attributed to massive algal blooms, which sank to the bottom consuming the available oxygen in the decay process.

Subsequent to the studies on Lakes Erie and Ontario, studies were conducted on Lakes Huron and Superior. These have been completed and recommendations have been made for controlling nutrient inputs because of identified and potential eutrophication problems (3).

In addition to problems in the Great Lakes, Ontario had, through previous studies, identified eutrophication as a serious problem and recognized the need for phosphorus control on the Ottawa River system and many inland recreational lakes (4).

A specific study on the Thames River watershed, initiated in 1972, identified that impairment of surface water quality was caused primarily by excessive inputs of nutrients, oxygen consuming materials (carbonaceous and nitrogenous), bacteria and suspended solids (5).

1.2 Concerns Elsewhere In Canada

In addition to the Great Lakes Region, several other areas in Canada have been identified, through watershed studies, as problem areas with concern for water degradation resulting from man's activities and specific to nutrient inputs (6).

1.2.1 Okanagan Basin - British Columbia

The Okanagan Basin, in the heart of British Columbia, is noted for its mild climate and beautiful lakes and has become a prime tourist and retirement area although it also has a reasonable industrial base.

For many years there were problems with water quantity and by the mid 1960's deterioration in water quality was evident. In response, detailed water management studies were carried out from 1969 to 1974.

The present quality of the lakes and streams was assessed and an inventory of all direct and indirect sources of wastes discharged into these waters was prepared. Studies were undertaken to determine the capacity of surface waters to assimilate present and future waste materials without adversely affecting potential uses.

Most streams in the basin had at least one constituent, such as colour, turbidity, phosphorus, nitrogen, or coliforms whose concentration exceeded acceptable health standards for drinking water. Varying degrees of nutrient enrichment, resulting in aquatic plant and algal growth, existed in all of the bottom valley lakes. The major sources of these nutrients were tributary streams and municipal outfalls, but septic tank systems and agricultural runoff were also

adversely affecting local shoreline areas.

Implementation of recommended control measures is to be undertaken between 1976 and 1981.

1.2.2 Qu'Appelle River Basin - Saskatchewan

The Qu'Appelle River Basin in eastern Saskatchewan and part of Manitoba encompasses an area of 51 800 square kilometres (20 000 square miles), with agriculture (wheat) being the most important industry. It is one of the major watersheds in southern Saskatchewan.

A water management study was carried out from August 1970 to March 1972 because of the importance of this river system to the region. The principal water quality problems in the Qu'Appelle basin were the highly eutrophic condition of the lakes, and the low dissolved oxygen levels in some parts of the system. To a large extent, these problems reflected the amount and nature of nutrients discharged into the system.

The concentrations of each of the major nutrients were found to be excessive, and adequate to support abundant aquatic plant growth.

The majority of urban centres in the basin with sewage treatment systems dispose of their effluent in tributaries of the Qu'Appelle River. Nearly 70% of the total amount of phosphorus and nitrogen in the water originates from municipal sewage systems; the City of Regina, the provincial capital, contributes 56%, and the city of Moose Jaw 13% of the total amounts. The remainder is derived from other sources such as eroded soils, agricultural wastes and groundwater.

Under the terms of the implementation agreement (1975 to

1985) wastewater treatment facilities were to be constructed at Moose Jaw, Regina, and other urban centres.

1.2.3 Saint John River Basin - New Brunswick

With a drainage area of 21,300 square miles, the Saint John River basin is one of the largest in the Atlantic Provinces. Slightly more than one-half of this area is located in New Brunswick, while the remainder lies in Maine and Quebec. The fisheries of the Saint John basin are an important, socially and economically, natural resource.

Although the main users of the water resources are forestry and forestry based industries, municipal uses of the surface water resource consist primarily of exploiting its properties of waste assimilation and transport.

In the Saint John River basin, as is common to all river basins in Canada, the development of new industry and the twin pressures of natural population increase and a general movement from rural to urban areas have altered water use patterns to the extent that constraints upon the water resources' availability and usefulness have begun to take on critical proportions.

A water management study, initiated in 1970 identified that improper municipal waste disposal is having an adverse effect on the basin fishery resources. Although waste contribution from municipalities do not compare in damaging effects to those of the food processing industry and pump mills, twenty-six of the 51 municipalities in the basin lack sewage treatment facilities.

The studies have shown that the basin has an abundant natural water resource, but that past lack of awareness has significantly reduced its usefulness. Careful management will be essential to the

preservation of these resources.

2.0 Identification Of The Critical Nutrient

Biological growth requires suitable climatic conditions and adequate chemical nutrients. These nutrients consist of organic and inorganic matter which fertilize the waters. The organic component in soluble form is food for bacteria, which if sufficiently fed, may dominate the biological system. When the nutrient input is inorganic, the primary producers (algae and rooted aquatics) are the major beneficiaries. This generation of organic matter provides food for predators, bacteria and scavengers thus the cycle repeats itself.

It is recognized that Carbon, Nitrogen and Phosphorus are the three major nutrients for biological production. A considered approach to the control of nutrients will examine the sources of these (and other) nutrients, recognize the limiting (or critical or essential or absolute) nutrient(s), and map out a practical course of action for the reduction or elimination of the input(s). In effect, the well-known "Law of the Minimum" applies, which states that variation in productivity is most often determined by limitations imposed by a lack of some nutritional element. In other words, as the least abundant nutritional element is used up, productivity decreases.

Inputs of all organic (carbonaceons) material provide the carbon source for photosynthesis and plant growth. However, limitation of this source is, obviously, not possible and practical. Moreover, a large percentage of aquatic plants can, and do, utilize the bicarbonate from the water itself to support photosynthesis and as most of the waters in these parts have high bicarbonate content, there

is no rationale in the control of this nutrient.

Nitrogen in combined inorganic forms such as nitrate, nitrite, or ammonium ion is required for algae growth. Although most algae can not make use of nitrogen gas, which is dissolved in natural waters in relatively large quantities, certain types, notably the blue-green algae, can fix gaseous nitrogen by converting it to ammonia and so make it available for their growth. Thus, where nitrogen is limiting, because of an over abundance of phosphorus, blue-green algae tend to flourish. These algae are especially likely to cause water quality related problems, such as tastes and odours in drinking waters and the curtailment of recreational activities due to masses of floating algae material.

Because of the ability of blue-greens to fix gaseous nitrogen, it would be very difficult, if not impossible, to control eutrophication problems through limitations on nitrogen inputs to a water body. There is an inexhaustible supply of gaseous nitrogen in the atmosphere and so, as the gaseous nitrogen in the water is used up by the nitrogen - fixing algae, it is readily replaced from the atmosphere. Also, since the nitrogen - fixing forms favored when nitrogen is limiting are especially prone to cause water quality problems, it is impractical and, in fact, undesirable to attempt to control eutrophication by restricting inputs of nitrogen even in areas where it is currently limiting to growth of most algae forms. To improve such areas, inputs of another major nutrient must be lowered to the point where it replaces nitrogen as the limiting factor and then lowered further so that the growth and yield of all algae forms is reduced.

Hence we turn to Phosphorus. During the years following World War II, major developments in the use of phosphorus resulted in

very significant increases in phosphate in waste discharges. The uses of phosphate detergents, and industrial discharges such as those from metal cleaning and rush-proofing operations have caused the concentration of phosphate in the raw wastewaters to rise four or five times above that found before the war.

For every 3 parts of phosphorus, theoretically, 100 parts of organic (carbon) material can be synthesized. Once this carbon becomes part of the biodynamic cycle, it behaves like any other carbon compound producing an oxygen demand of 2.68 times its weight after the decay of the algae and providing a carbon nutrient source for bacteria. So, if we consider phosphorus as the critical element, we find that its removal should decrease the phosphorus/nitrogen ratio, and the green algae should so reduce the concentration of phosphorus in the aquatic environment that growth of nitrogen-fixing blue-green algae will be impossible. The important feature in the control of nutrient pollution is that phosphorus offers the best combination of a nutrient that is both growth-limiting in the lakes and at the same time controllable by man.

3.0 Sources of Phosphorus

Compounds of phosphorus are widely distributed throughout the earth. Phosphorus containing compounds are relatively scarce in the atmosphere, but a virtually unlimited quantity (10^{19} tons) of phosphorus exists in the earth's solid sphere (7). Fortunately most phosphorus containing minerals are only very slightly soluble in water and this tends to limit the concentrations found in natural waters to very low levels. Excessive levels of phosphorus in waters can almost assuredly be traced to man's activities.

The sources of phosphorus which are of most significance to eutrophication are many and diverse as listed in Table 1. Although the significance of each source to the fertilization of a specific waterbody is very site dependent, in the highly populated areas of the Lower Great Lakes, municipal wastewater treatment plant discharges are of greatest significance.

The 1970 report to the International Joint Commission (IJC) based on a 6-year water quality study of the Lower Great Lakes identified the sources of phosphorus in 1967 to Lake Erie and Ontario as presented in Table 2(1). Municipal sewage treatment plant effluents accounted for 63% of the total phosphorus load to Lake Erie and 54% to Lake Ontario.

Phosphorus in municipal sewage originates from human wastes, food wastes, industrial discharges to the sewer and phosphorus based detergents. Out of an annual per capita contribution of about 3.5 pounds of phosphorus in domestic sewage, Sawyer (8) estimated that 50 to 75% originated from phosphorus based detergents. The above IJC study, estimated that of the total phosphorus entering Lake Erie from municipal sources, 74% originated from detergents. The corresponding figure was 60% for Lake Ontario, thus singling out detergents as by far the major single source of phosphorus to the Lower Great Lakes.

Although this report identified municipal effluent as the major source of phosphorus to the Lower Great Lakes, it also concluded that the then current conditions in the lakes could not be related entirely to pollutant loadings from readily definable point sources. Studies indicated that 30 and 43 percent of the total phosphorus loads for Lakes Erie and Ontario, respectively, were due to sources other than municipal sewage and industrial effluents.

TABLE I

SOURCES OF PHOSPHORUS TO LAKE WATERS

I. Wastes from municipal treatment plants and private waste disposal systems

- A. Domestic
  - 1. Human wastes
  - 2. Spent soap and detergent waste waters
  - 3. Household food including garbage disposal wastes
- B. Industrial wastes discharged into municipal treatment plants
- C. Other

II. Industrial wastes discharged directly to waterways

III. Water treatment chemicals

IV. Land runoff

- A. Urban runoff and drainage
- B. Storm water
- C. Rural runoff and drainage
  - 1. Agricultural
    - a. Soils
    - b. Irrigation water
    - c. Fertilizers and manure
    - d. Farm animal and bird wastes
  - 2. Non-agricultural land
    - a. Decaying leaves and other decaying vegetation from woodlands
    - b. Wild animal and bird wastes

V. Ground water as a lake water source

VI. Reserves in lakes

- A. Lake sediments ("bottom muds")
- B. Living aquatic plant and animal organisms
- C. Lake water

VII. Atmosphere

A. Rainfall

B. Dustfall

TABLE 2

Input of Total Phosphorus in 1967 (1)

Source	Metric Tons/Year To Lake Erie	To Lake Ontario
Lake Huron	2009	--
Lake Erie	--	4018
Municipal	17054	6562
Industrial	1786	402
Land Drainage	6027	714
Unaccounted	--	491
Total	26,876	12,187

The International Reference Group on Great Lakes Pollution from Land Use Activities (PLUARG) was later to further define non-point sources of pollution (9). PLUARG confirmed that phosphorus loads from land drainage and atmosphere deposition did indeed contribute to both offshore and nearshore water quality problems related to eutrophication. Depending upon the magnitude of the point source loads, PLUARG estimated that the combined land drainage and atmospheric inputs to individual Great Lakes ranged from 32 percent (Lake Ontario) to 90 percent (Lake Superior) of the total phosphorus loads excluding shoreline erosion. Intensive agricultural operations were identified as the major diffuse source contributor of phosphorus. Excessive fertilizer applications and improper manure management were considered the major causes.

Runoff from urban areas arising from sanitary sewer, combined sewer and storm sewer overflows; industrial discharges; street and ground surface runoff and other non-point sources have been identified as another as yet unquantified source of phosphorus to receiving waters (10).

Certain sources of phosphorus such as point source municipal and industrial wastewater treatment plant effluents are fairly readily controllable. Other diffuse sources such as agricultural runoff are difficult to define and difficult to control. Therefore, a phosphorus control program must evaluate the relative significance of all sources and predict the practicality and resultant impact of the control of each source before identifying a specific control strategy.

4.0 Ontario's Approach To Phosphorus Control

Arising out of the 6-year study of pollution in the Lower Great Lakes drainage basin, the International Joint Commission (IJC) in 1969 recommended that phosphorus discharges from all sources in the Lower Great Lakes be reduced to the lowest practical level (11).

Recognizing the significance and controllability of municipal wastewater treatment plant discharges as a major source of phosphorus, the Province of Ontario, in 1970, announced a policy requiring the installation of phosphorus removal facilities at municipal and institutional wastewater treatment plants located not only in the Lower Great Lakes but also in the Ottawa River system, parts of the Upper Great Lakes and throughout the inland recreational areas. Previous studies by the Ministry of the Environment (formerly the Ontario Water Resources Commission) on waters in the latter areas had shown phosphorus to be a major factor influencing eutrophication.

Initially, the policy required a minimum removal of 80% of the phosphorus from the wastewater plant influent with higher levels of removal to be determined by further studies of the receiving waters. This criterion was subsequently superceded in the Lower Great Lakes by the signing in April, 1972 of the Canada-United States International Agreement on Great Lakes Water Quality (12) which called for an effluent objective of 1 mg/l total phosphorus.

Under this agreement, permanent phosphorus removal facilities were to be operational by December 31, 1973 in the most critically affected areas of the Province; by December 31, 1975 for those plants discharging to waters deemed to be in a less critical condition; and three years after notification in all other areas of the Province where problems are found to exist (See Figure 1). The 1973 phase of the

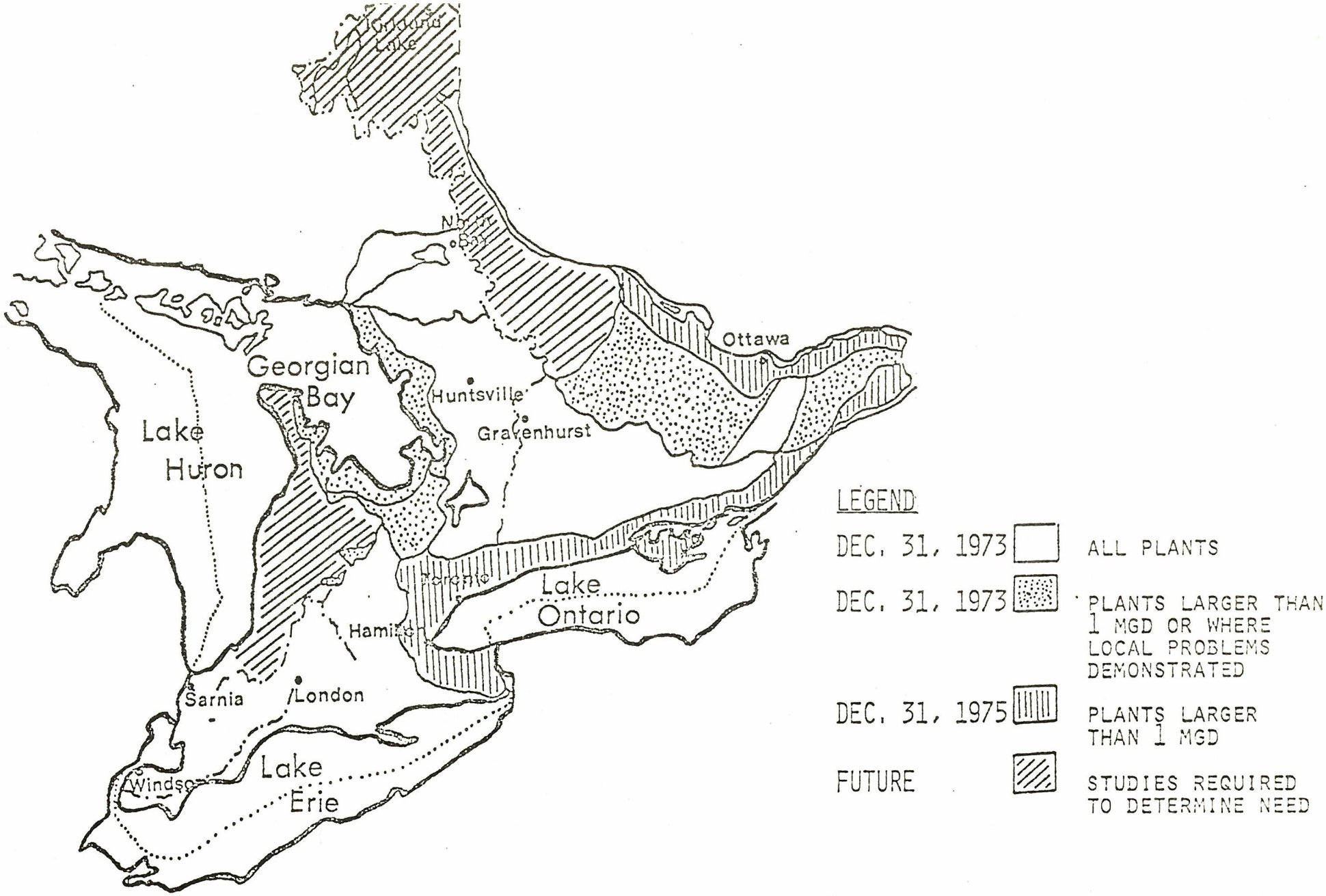


FIGURE 1. PROVINCE OF ONTARIO - SOUTHERN SECTION
PHOSPHORUS REMOVAL PROGRAM SCHEDULED COMPLIANCE DATES (4)

program involved approximately 110 plants with capacities ranging from 0.04 to 24 IMGD, serving an estimated 1,600,000 population. The 1975 phase brought into operation another 50 plants ranging in size from 0.3 to 180 IMGD, serving an additional 3,100,000 persons.

Concurrently, the Government of Canada, under terms of the Canada Water Act (1970), enacted a policy calling for a staged reduction in the phosphorus content of detergents to a limit of 8.7% phosphorus (as P) by weight by August 1970 and 2.2% by December 31, 1972.

This action recognized that detergents were the major contributor of phosphorus to municipal wastewaters and that since some time would necessarily be required before municipal treatment plants could install phosphorus removal processes, such action would have an immediate effect on the discharge of phosphorus from these sources. Such detergent reformulation regulations would also have immediate and lasting impact in those areas of the country which did not have treatment facilities.

As a prelude to the Canada-United States Agreement, the Governments of Canada and Ontario, in August 1971, signed an agreement (13) to implement and accelerate programs of pollution control in the Lower Great Lakes to meet the recommendations of the IJC. The agreement secured funding for a $250 million capital works program aimed at upgrading sewage collection systems and treatment works, including the installation of phosphorus removal equipment at an estimated cost of $40 million. An additional $6 million over the 5-year term of the agreement was provided for related research studies. Phosphorus removal technology and treatability studies were to be given top priority in the research program. The $6 million research fund was jointly shared and administered by the Governments of Canada and

Ontario. The agreement was subsequently extended for an additional 2-year period with a further $1 million allocation of research funding.

Largely as a result of this research funding a program of integration of chemical treatment into existing plant processes involving jar testing and treatability studies at each plant was established. Consequently the deadlines for phosphorus removal implementation established by the province in 1970 were largely met.

In spite of the concerted effort that Ontario has placed on reducing phosphorus from municipal sewage treatment plant effluents, other sources are not being ignored. Specifically: 1) Ontario policy calls for an effluent objective for industrial phosphorus discharges of 1 mg/l maximum total phosphorus. 2) The Urban Drainage Sub-Committee of the Canada/Ontario Agreement in their Proposed Model Policies for Urban Drainage Management (10) suggests that "each municipality shall formulate and implement a comprehensive pollution control strategy related to its own particular land use, drainage and runoff characteristics". This strategy should not only consider sewage treatment plant effluent but also sanitary sewer, combined sewer and storm sewer bypassing, industrial discharges and non-point drainage sources. 3) The province has recently adopted sewage sludge utilization guidelines (14) which include provisions for phosphorus control. 4) PLUARG in its summary report to the IJC (9) recommends that phosphorus reduction through non-point as well as point source control programs be implemented. This would include control of soil erosion, strengthening of the Ontario Agricultural Code of Practice (15) in the areas of reducing water pollution problems from animal manures, and improving fertilizer application practices.

5.0 Impact Of Ontario's Phosphorus Control Program

As discussed previously, the approach to phosphorus control in Ontario was basically a two-prong attack; reduction in the phosphorus content of detergents and removal of phosphorus at municipal (and industrial) wastewater treatment facilities. The combined effect of these two measures has been quite dramatic.

5.1 Effect of Detergent Regulations

Prior to the Canadian Government imposing any controls on detergent phosphorus levels, the phosphorus content of laundry detergents averaged out at 11%. The raw sewage phosphorus concentration correspondingly averaged about 10 to 11 mg/l.

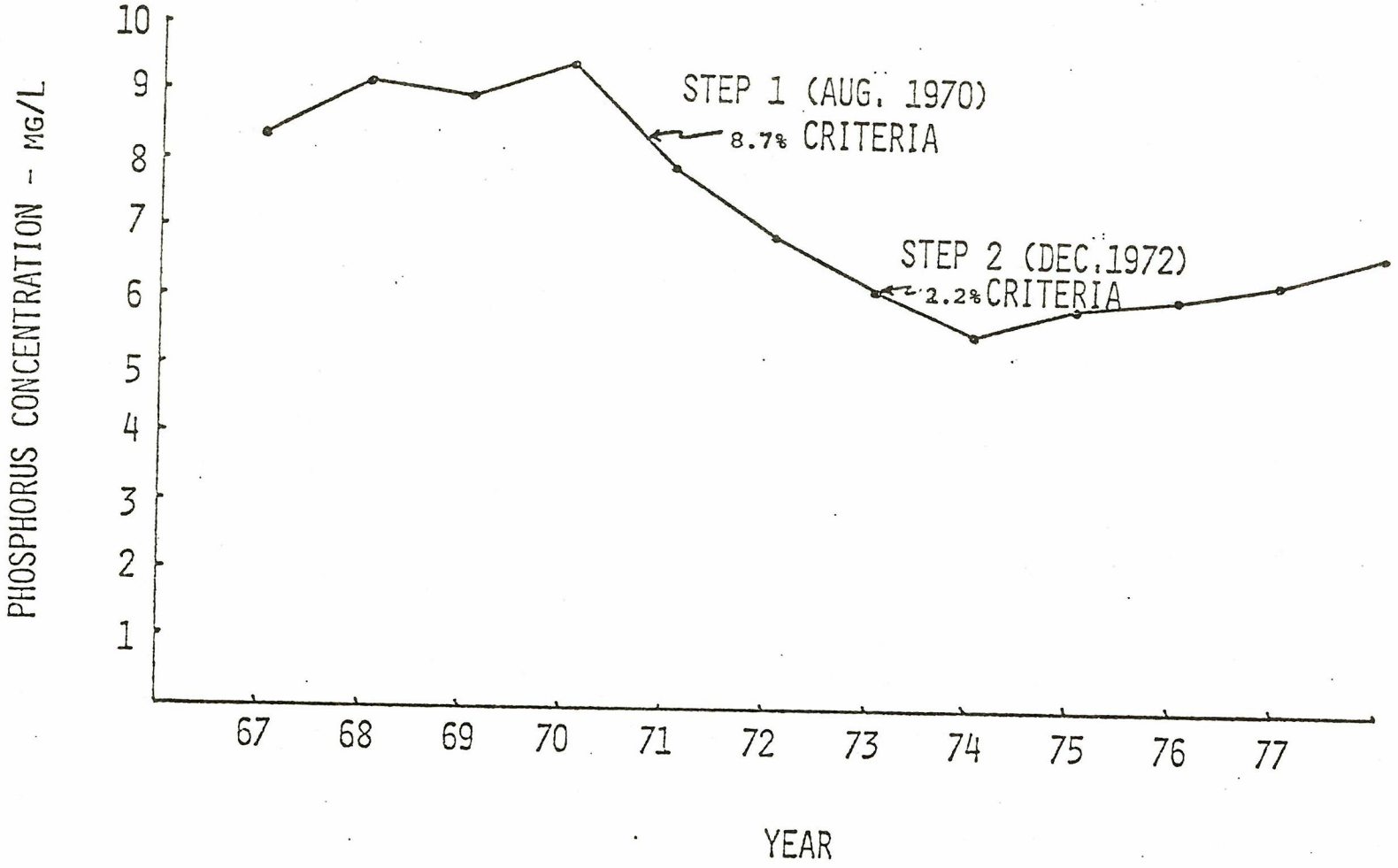
Following the introduction of the first control regulations effective August 1970, which limited laundry detergent phosphorus to 8.7%, the raw sewage concentration decreased to about 8.8 mg/l. (The actual phosphorus content of detergents for the 2½ years that this regulation was in effect averaged 8.4% (16)).

After the control regulations of January 1, 1973 were introduced which further limited detergent phosphorus to 2.2%, the raw sewage phosphorus concentration dropped to about 7 mg/l. (The actual detergent phosphorus content averaged 1.9%).

This raw sewage concentration continued to fall through 1974 to a low value of about 5.6 mg/l but has since increased to a current level of about 6.8 mg/l (17). Figure 2 relates raw sewage phosphorus concentration to detergent regulations for the years 1969 to 1978. In regions which do not have a detergent phosphorus control regulation in effect, approximately 35% of the phosphorus in domestic sewage is derived from phosphate based detergents (18).

FIGURE 2

PHOSPHORUS CONCENTRATIONS IN RAW SEWAGE



This value is considerably below the figures developed by Sawyer (8) and Vollenveider (19), which indicated that over 50% of domestic sewage phosphorus was derived from detergents. At that time (1968) however, detergent phosphorus content was typically 11% as compared to the current level in unregulated areas of about 5%.

In effect, the detergent phosphorus control regulation then, has reduced the phosphorus concentration in raw sewage by about 35%. This could be translated into a reduction of effluent phosphorus concentrations from a primary plant without phosphorus removal from 8.4 mg/l to 5.4 mg/l, and from a secondary plant without phosphorus removal from 4.2 mg/l to 2.7 mg/l. In actuality, detergent reformulation caused an overall reduction in effluent phosphorus concentration in all Ontario plants, including both primary and secondary facilities, from an average of 5.5 mg/l in 1969 to 3.5 mg/l in 1973 before an appreciable number of phosphorus removal facilities were in operation (17).

In areas which do not have phosphorus removal facilities, this reduction could significantly reduce receiving water phosphorus loadings. In areas with phosphorus removal this reduction means a marginal reduction in chemical requirements.

5.2 Phosphorus Removal At Municipal WPCP's

Besides the federal government imposing household detergent reformulation requirements, the Province of Ontario implemented a policy requiring the installation of phosphorus removal facilities at communal and institutional water pollution control plants (WPCP's) not only in the Lower Great Lakes Basin, but also in parts of the Upper Great Lakes, the Ottawa River system and the inland recreational

lakes area.

Table 3 presents the total number of wastewater treatment plants with operational phosphorus removal facilities as of 1973, 1975 and 1977. The 211 plants represent about 85% of provincial total hydraulic capacity.

Phosphorus removal in Ontario involves a chemical precipitation process in which aluminum or iron salts are used as the prime coagulant within the sewage treatment plant process. Lime is also being utilized as an alternative in a small number of locations. Along with their ability to precipitate phosphorus within the conventional sewage treatment processes, the coagulants tend also to remove greater quantities of heavy metals from solution and in primary plants considerably improve BOD and Suspended Solids removals.

The 1978 Annual Report on Great Lakes Water Quality (20) assesses the phosphorus removal programs for 1978. As a result of the programs, municipal phosphorus loadings have decreased in the Lake Erie and Lake Ontario basins. Ontario loads were below the target level for Lake Erie but slightly above that for Lake Ontario. The United States loads are still more than double the target loads for both lake basins. The phosphorus effluent target of 1.0 mg/l or less was achieved in 1978 at 40 of 68 Ontario plants but only 32 of 112 U.S. plants reported as discharging more than 1 MGD in the Lakes Erie and Ontario basins.

7.0 Nitrogen Control In Ontario

Since this paper addresses the topic of nutrient control further word must be said about nitrogen.

Nitrogenous compounds discharged from wastewater treatment

TABLE 3

PHOSPHORUS REMOVAL FACILITIES WITHIN THE PROVINCE OF ONTARIO

Watershed	Total Number of Treatment Plants With Phosphorus Removal		
	1973	1975	1977
Lake Superior	0	0	0
Lake Huron	33	31	37
Ottawa River		66	80
Lake Erie	78	62	77
Lake Ontario		9	17
St. Lawrence			
Total	111	168	211

plants can have several deleterious effects. Although biostimulation of receiving waters is an important role of nitrogen we have already concluded that there would be no benefit from controlling nitrogen discharges for that purpose. Other less apparent impacts of nitrogen can be of major importance in specific situations. These impacts include toxicity to fish life, reduction of chlorine disinfection efficiency, an increase in the dissolved oxygen depletion of receiving waters, adverse public health effects and a reduction in the suitability of the wastewater for reuse. In Ontario concerns over nitrogenous compounds in sewage effluents are limited to fish toxicity and oxygen depletion.

The principal toxicity problem is from ammonia, in the molecular form (NH_3), which can adversely affect fish life in receiving waters. A slight increase in pH may cause a great increase in toxicity as the ammonium ion (NH^{+4}) is transformed to ammonia. Factors which may increase ammonia toxicity at a given pH are greater concentrations of dissolved oxygen and carbon dioxide, elevated temperature, and high bicarbonate alkalinity. Reported levels at which acute toxicity is detectable have ranged from 0.01 mg/l to over 2.0 mg/l of molecular ammonia nitrogen (21). Ammonia toxicity is best eliminated by reducing the effluent ammonia concentration to very low levels (i.e. under 1 mg/l).

In receiving waters, ammonia can be biologically oxidized to nitrite and then to nitrate, thereby adding to the oxygen depletion caused by carbonaceons materials. Ammonia has an oxygen demand of 4.6 mg O_2 per mg NH_3 and therefore a typical secondary effluent with 15 mg/l BOD and 15 mg/l NH_3 will have a total oxygen demand of 84 mg/l. This high level of oxygen demand may cause significant oxygen depletion in the receiving water if insufficient dilution is available.

As we progress to increasingly efficient BOD removal processes, effluent ammonia concentration can no longer be ignored.

Just as ammonia may be oxidized to nitrite and then nitrate in a receiving water, so can the activated sludge process be modified to effect ammonia oxidation, a process termed nitrification. Several municipal sewage treatment plants in Ontario, situated on critical receiving waters are now required to include nitrification processes. No nitrogen removal, per se, is required in Ontario.

8.0 Costs of Nutrient Control in Ontario

Since the phosphorus removal program was integrated into a $480 million capital works program for the construction of new and expanded sewage treatment plants, trunk sewers and collection systems it is virtually impossible to determine the exact costs of Ontario's phosphorus removal program. Original projections of capital costs at the policy formulation stage were in the neighbourhood of $40 million.

The Ministry of the Environment estimated the actual capital cost for equipment purchase and installation for the phosphorus removal program at $15 million (22). The figure has since been revised down to $13.5 million and is exclusive of any plant enlargements or expansion of sludge handling facilities. This works out to a capital cost of about $16,000 per MG plant capacity. The savings of $20.5 million over the projected cost has been attributed to the related research work and the development of the treatability study procedure.

Operating costs for phosphorus removal are almost equally as difficult to define and are very site and size dependent. Average costs however have been determined as $23 per MG treated for phosphorus removal chemical and $380 per year operating expense exclusive of

chemical costs for each MG of design capacity (23).

Costs of nitrification are too site specific for comment.

9.0 Summary

In summary, Ontario's nutrient control program has been to reduce phosphorus discharge levels in municipal and industrial wastewater treatment plant effluents and to reduce effluent ammonia nitrogen levels in specific situations, while keeping in mind the contribution from other sources.

Phosphorus reduction has been achieved quite creditably through the federal government's detergent reformulation regulations and through chemical precipitation processes at operating wastewater treatment facilities.

Ammonia nitrogen reduction is being effected through the conversion of identified wastewater treatment facilities to combined carbonaceons removal/nitrification processes.

Costs of phosphorus removal have been significant but were much less than were predicted at the initiation of the program.

References

1. International Joint Commission, "Pollution of Lake Erie, Lake Ontario and the International Section of the St. Lawrence River" (1970).

2. Burns, N. and C. Ross, "Discussion of Findings of Project Hypo", in "Project Hypo", CCIW Paper No. 6, U.S. EPA Technical Report 5-05-71-208-24, (1971).

3. Upper Lakes Reference Group, "The Waters of Lake Huron and Lake Superior, Volume I, Summary and Recommendations", report to the IJC by the Upper Lakes Reference Group, (1976).

4. Van Fleet, G.L. "Phosphorus Removal in Ontario", Phosphorus Removal Design Seminar, Canada-Ontario Agreement Conference Proceedings No. 1, (1973).

5. Salbach, S.E., "Thames River Basin Water Management Study", High Quality Effluents Seminar, Canada-Ontario Agreement Conference Proceedings No. 3, (1976).

6. Canada Water Year Book, Environmental Management Service, Fisheries and Environment Canada, (1976).

7. Van Wazer, J.R. "Phosphorus and Its Compounds, Vol. 11, Technology, Biological Functions, and Applications" Interscience Publishers Inc., New York (1961).

8. Sawyer, C.N. "Some New Aspects of Phosphates in Relation to Lake Fertilization". Sewage and Industrial Wastes 24 (1952).

9. International Reference Group on Great Lakes Pollution From Land Use Activities (PLUARG), "Environmental Management Strategy for the Great Lakes System". Final Report to the International Joint Commission (July 1978).

10. Urban Drainage Subcommittee of The Canada-Ontario Agreement on Great Lakes Water Quality, "Manual of Practice on Urban Drainage".

11. International Joint Commission, "Pollution of Lake Erie, Lake Ontario and the International Section of the St. Lawrence River", Vol. 1, Summary (1969).

12. Canada-United States International Agreement on Great Lakes Water Quality (April, 1972).

13. Canada-Ontario Agreement on the Lower Great Lakes (August, 1971).

14. Ontario Ministry of Agriculture and Food-Ontario Ministry of The Environment, "Guidelines for Sewage Sludge Utilization on Agricultural Lands" (April, 1978).

15. Ministry of The Environment and The Ministry of Agriculture and Food, "Agricultural Code of Practice for Ontario (April, 1973).

16. Appleby, D.J., "The Impact of Phosphorus Control Activities-The Experience in Ontario". Erco Industries Ltd., (October, 1977).

17. Ministry of The Environment, "Operating Summary-Water Pollution Control Projects" (1978).

18. Lee, G.F., and Jones, R. Anne, "Effect of Eutrophication on Fisheries". Department of Civil Engineering, Colorado State University, (1979).

19. Vollenweider, R.A. "Scientific Fundamentals of the Eutrophication of Lakes and Flowing Waters, With Particular Reference to Nitrogen and Phosphorus as Factors in Eutrophication". Tech. Report DAS/CSl/68.27, OECD, Paris, (1968).

20. International Joint Commission, "Great Lakes Water Quality 1978 Annual Report", (July, 1979).

21. AWWA Task Group Report. "Sources of Nitrogen and Phosphorus in Water Supplies", Journal AWWA, (March, 1967).

22. Archer, J.D. "Summary Report on the Phosphorus Removal Program", Ontario Ministry of The Environment publication, (April, 1976).

23. Archer, J.D., "Summary Report on Phosphorus Removal", Canada-Ontario Agreement Research Report No. 83, (1978).

PHYSICAL-CHEMICAL PHOSPHORUS REMOVAL PROCESSES

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INTRODUCTION

In 1969, the International Joint Commission (IJC) recommended that phosphorus discharges from all sources to the Great Lakes be reduced to the lowest practical level. Concurrent with the IJC recommendation the Province of Ontario established a policy requiring phosphorus removal at municipal and institutional wastewater treatment plants discharging to Lakes Erie, Ontario or inland recreational waters.

In 1972 the IJC recommendation was further supported by the signing of the Canada-U.S.A. Agreement on Great Lakes water quality which called for an effluent objective of 1 mg/L total phosphorus. In order to implement the Canada-U.S.A. agreement mandate the governments of Canada and Ontario signed a five year agreement which provided $250 million of capital works funding for phosphorus removal systems and an additional $6 million earmarked for research and development of cost-effective phosphorus removal systems. The research agreement was subsequently extended for two years with an additional funding of $1 million.

Over the seven year period (1971-1978) that the Canada-Ontario Agreement was in effect extensive bench, pilot and full scale investigations were carried out on virtually every aspect of phosphorus removal. Much of the total research effort of the Canada-Ontario Agreement was conducted and administered by the Ministry of Environment, Ontario and the Environmental Protection Service of Environment Canada. This paper relies heavily on the experience and knowledge gained from the Canada-Ontario research effort to illustrate various aspects of chemical phosphorus removal.

PHOSPHORUS IN WASTEWATER

Phosphorus is an essential element in the metabolism of organic matter. Consequently a certain amount is required in order to adequately treat wastewater biologically.

The total phosphorus content of municipal wastewater is comprised of (1) suspended phosphorus (either organic phosphorus bound up in cell protoplasm or phosphorus adsorbed onto particulate matter), (2) Condensed phosphorus (inorganic polyphosphates originating mainly from detergents) and (3) ortho phosphates (largely soluble and representing the final breakdown product of the phosphorus

cycle). The relative magnitude of each form in typical domestic wastewater with and without high phosphate detergents is shown in Figure 1. In Ontario detergent reformulation programs have reduced the total phosphorus content of municipal wastewates from an average of 7.0 mg/L in 1969 to 5.2 mg/L in 1974.

When a municipal wastewater is subjected to conventional wastewater treatment (i.e., primary plus secondary biological processes) about 0.5 to 1.0 grams of phosphorus will be removed for each 100 grams of BOD_5 reduction. This results in an overall phosphorus removal through a typical plant of 20 to 40%. The fate of phosphorus within a secondary plant is shown in Figure 2. Poly phosphates are readily hydrolized to ortho-phosphates by biological systems and as a result the secondary effluent total phosphorus content will consist largely of the simpler ortho-phosphates.

Although biological systems are capable of being operated to produce enhanced phosphorus removal (e.g., the Bardenpho process developed in South Africa and the Phostrip system developed by Union Carbide) it is questionable that they can consistently produce an effluent total phosphorus equal to or less than 1 mg/L. Since 1 mg/L or less is the usual effluent removal criterion, essentially all full-scale phosphorous removal systems have been based on the more effective and reliable chemical precipitation processes.

PHOSPHORUS REMOVAL CHEMICALS

Several chemicals or chemical wasteproducts have been utilized for phosphorus removal. In general these are ionic forms of aluminum, iron or calcium. Table 1 lists several of the more common phosphorus precipitants.

TABLE 1

PHOSPHORUS PRECIPITANTS

Aluminum Sulphate (Alum)	$Al_2(SO_4)_3 \cdot 14H_2O$
Iron Salts	Fe^{3+} Fe^{2+} Chlorides or Sulphates
Hydrated Lime	$Ca(OH)_2$
Sodium Aluminate	$Na_2O \cdot Al_2O_3$
Pickle Liquors	High Fe^{3+} Content
Waste Alumium Chloride	$AlCl_3$
Carbide Lime	From Acetylene Production

The basic removal mechanism for all of these chemicals is due to the formation of an insoluble precipitate comprised of the metallic ion and orthophosphate. In addition condensed and suspended phosphorus is removed by a combination of complex reactions or adsorption onto the flocculant precipitate.

In practice phosphorus removal is achieved within existing wastewater systems by using alum, iron salts or lime in a primary precipitation mode (Figure 3) or alum or iron salts in a secondary mode (Figure 3). For more stringent effluent phosphorus requirements (i.e., 0.5 mg/L or less) post precipitation (Figure 4) with alum, iron salts or lime may be utilized. The basic mechanism and major factors influencing phosphorus removal by alum, iron salts and lime are discussed in the following sections.

Aluminum

When alum is added to wastewater containing phosphate the following idealized reaction occurs:

doesn't work

$$Al_2(SO_4)_3 \cdot 14\,H_2O + 2PO_4^{3-} \longrightarrow \underset{\downarrow}{2AlPO_4} + 3SO_4^{2-} + 14H_2O$$

The resultant precipitate of aluminum phosphate can be separated by conventional solid-liquid separation processes such as sedimentation.

Alum also will react with the wastewater alkalinity as follows:

$$Al_2(SO_4)_3 \cdot 1AH_2O + 6HCO_3^- \longrightarrow 2Al(OH)_3 + 6CO_2 + 14H_2O + 3SO_4^{2-}$$

This reaction will tend to lower the pH of the wastewater. The optimum pH for aluminum phosphate precipitation is around 6; therefore the wastewater must have enough alkalinity to buffer the alum. In exceptionally few cases of low wastewater alkalinity some supplemental alkalinity is required to maintain a favourable pH.

Iron

When iron salts are added to wastewaters containing phosphate the following idealized reaction occurs:

$$Fe^{3+} + PO_4^{3-} \longrightarrow \underset{\downarrow}{FePO_4}$$

As indicated in the reaction and as demonstrated in numerous phosphorus removal investigations the iron must be in the ferric form before significant phosphorus reduction occurs. However ferrous iron once oxidized to the ferric state:

$$4Fe^{2+} + O_2 + 4H^+ \longrightarrow 4Fe^{3+} + 2H_2O$$

will achieve effective phosphorus removal. It has been determined that approximately 0.15 g of oxygen are required per g of Fe^{2+} for the conversion.

Calcium

Calcium reacts with phosphate ions in the presence of hydroxl irons to form an insoluble precipitate hydroxyapatite.

$$5Ca^{2+} + 3HPO_4^{2-} + 4OH^- \longrightarrow \underset{\downarrow}{Ca_5(OH)(PO_4)_3} + 3H_2O$$

Lime also reacts with the natural bicarbonate alkalinity and free carbon dioxide of a wastewater to precipitate calcium carbonate. Only the excess calcium remaining from this reaction is available to precipitate phosphate. In general, the quantity of lime required for phosphorus removal will be independent of the amount of phosphorus and directly related to wastewater alkalinity.

Phosphorus removal with lime can be achieved by the "so-called" low-lime process (pH 9.5 to 10) and the high-lime process (pH > 11). The former process, under optimum conditions, is capable of producing an effluent with less than 1 mg/L of total phosphorus whereas the latter process can produce effluents of 0.5 mg/L or less of total phosphorus. The high-lime process is discussed in a later section.

CHOICE OF PRECIPITANT

Basically, the designer of a phosphorus removal system whether it be a retrofit for an existing system or a new system has a choice of three chemical precipitants, i.e., alum, iron salts or lime. In order to select the most cost effective chemical for a particular application treatability testing and cost comparisons are necessary. In Ontario a two-phase treatability approach has been an integral part of the overall phosphorus removal program. Jar testing and full-scale treatability testing has been carried out at over 75 separate treatment plants in Ontario. This experience has resulted in a large data base

from which predictive methodologies have been developed. It is now felt that the full-scale treatability testing need not be carried out unless peculiar wastewater characteristics, plant design or operational conditions prevail.

Jar Testing

The Ontario Ministry of Environment (MOE, 1972) has developed a detailed procedure for jar testing which utilizes standardized rapid mixing, slow mixing and quescent settling times.

The preliminary testing is done on batch raw sewage and final effluent grab samples using the three primary coagulants, aluminum sulfate, ferric chloride and lime at a wide range of dosages. Additional jar tests are subsequently conducted to narrow down the range of coagulant dosage needed to effect the required degree of phosphorus removal, expressed either as a percentage removal or a residual phosphorus concentration.

Finally, a number of jar tests are conducted using dosages that attempt to bracket the range of removal efficiency desired, i.e., dosages yielding removals lower, equal to and higher than the objective. It is essential that these jar tests be conducted over an extended period of time in order that a representative variety of sewage characteristics are encountered. Grab samples for jar testing should be taken at different times of the day and on various days of the week. Once the range of chemical dosages has been determined, it is essential not to alter these dosages used in the jar testing in order to avoid deliberately skewing the results. In order to have sufficient data to draw meaningful conclusions, it was arbitrarily considered that a minimum of ten data points should be obtained for each chemical dosage on each waste stream being studied.

Proper analysis of the data obtained from jar testing is essential in obtaining meaningful information on the relative efficiencies of each chemical and point of addition. Rather than looking at the average total phosphorus removal for each set of data, plots of each set of data points are made on arithmetic probability paper. Such probability curves have two significant aspects: the relative vertical position of the point indicates the degree of effectiveness of the coagulant used at that specific dosage; and, the slope of the line indicates the expected reliability of phosphorus removal using that particular coagulant.

A typical set of data using alum as the coagulant for raw sewage is shown in Figure 5. This figure indicates that in order to achieve a residual phosphorus concentration of 1 mg/L, 50 per cent of the time, a dosage of 225 mg/L is probably adequate. A dosage of 300 mg/L, however, is required to achieve 1 mg/L 80 per cent of the time.

Using similar curves for each coagulant, the dosage required to yield a given degree of phosphorus removal (or residual phosphorus value) for a specified frequency may be determined and results compared.

The relative economics of the chemicals and dosage points for phosphorus removal can then be determined on the basis of optimum dosage, delivered cost and availability of chemical.

Full-Scale Treatability Tests

Full-scale temporary treatability studies were conducted using the optimum chemical selected by jar testing. Full-scale studies were necessary, initially, to confirm chemical dosage, optimize the point of application; provide data on sludge production and dewaterability; determine expected final effluent quality and provide operational cost data.

For primary sewage treatment plants, full-scale temporary studies should be of approximately six weeks' duration. Eight week full-scale studies are considered necessary for secondary treatment facilities. The above study durations assume that jar tests have been used to establish the optimum chemical and general point of application (i.e., to raw sewage or secondary effluent) and that the chemical selected and its point of application yield satisfactory results. If not, the study duration must be expanded accordingly. Additionally, the presence of a significant industrial waste load may necessitate more extensive studies in both the jar testing and full-scale work.

The primary objective of the full-scale temporary phosphorus removal study is to demonstrate the feasibility of achieving a specific phosphorus removal objective at a particular sewage treatment plant without requiring extensive plant modifications or addition.

Further details of the treatability studies carried out in Ontario are presented in a report by Boyko and Rupke (1976). A summary of the treatability data has been compiled by Prested et al (1977) and and selected data are presented in Table 2.

TABLE 2

BASIC STATISTICS FROM THE CANADA-ONTARIO
FULL-SCALE TREATABILITY STUDY PROGRAM

CHEMICAL	TOTAL NO. OF PLANTS	ADDITION TO RAW WASTEWATER			ADDITION TO MIXED LIQUOR		
		NUMBER OF PLANTS	AVG. DOSE (mg/L)	AVG. COST[3] ($/MILLION GALLONS)	NUMBER OF PLANTS	AVG. DOSE (mg/L)	AVG. COST[3] ($/MILLION GALLONS)
Iron	22	7	16.0[1]	40.3	15	9.2	22.5
Alum	20	5	10.3[1]	67.0	15	7.2	43.4
Lime	3	3	118[2]	35.0	-	-	-

[1] - Dosages as mg/L Fe/Al.

[2] - Dosages as mg/L Ca.

[3] - Chemical cost only (adjusted to December 1979).

The optimum method for phosphorus removal in most instances has involved the addition of ferric chloride or alum to the mixed liquor (i.e., 30 out of 45 plants). Of the twelve plants that tested full-scale addition of alum or ferric chloride to the raw wastewater seven were primary plants and had no other system. Only three plants tested lime addition on a full-scale.

The chemical cost for phosphorus removal based on the average full-scale chemical dosage, wastewater flow and actual delivered chemical cost (adjusted to December, 1979 prices) were calculated by the consultants for each full-scale study. On the basis of the average chemical costs, ferric chloride addition to the mixed liquor was the most economical method of meeting an effluent phosphorus concentration of 1 mg/L when retrofitting an existing secondary plant. However, it must be realized that delivered chemical costs will influence each particular situation and that data presented herein are for a geographically diverse group of Ontario municipalities.

Statistical Prediction of Chemical Dosage

Prested et al (1977) have taken the Ontario treatability data, standardized them and carried out regression analyses to develop predictive equations for chemical phosphorus removal. Some of these equations are summarized in Table 3.

TABLE 3: REGRESSION RELATIONSHIPS BETWEEN CHEMICAL REQUIREMENTS AND INFLUENT WASTEWATER PHOSPHORUS LEVELS

Chemical	Jar Test Data**		Full Scale Data**	
	Raw Wastewater Addition	Mixed Liquor Addition	Raw Wastewater Addition	Mixed Liquor Addition
Iron* (mg/l as Fe^{3+})	Fe^{3+} = 2.8+3.3(TP) r = .653 n = 71 Ave Fe^{3+} = 24.3 Ave TP = 6.5 Ave Fe^{3+}/TP = 3.7	Fe^{3+} = 2.4+3.8(TP) r = .650 n = 57 Ave Fe^{3+} = 17.8 Ave TP = 4.0 Ave Fe^{3+}/TP = 4.5	Insufficient data for regression n = 7 Ave Fe^{3+} = 16.0 Ave TP = 5.3 Ave Fe^{3+}/TP = 3.0	Fe^{3+} = .5+1.4(TP) r = .689 n = 15 Ave Fe^{3+} = 9.2 Ave TP = 6.0 Ave Fe^{3+}/TP = 1.5 Ave Residual TP = .9
Alum (mg/l as Al^{3+})	Al^{3+} = .6+2.4(TP) r = .689 n = 72 Ave Al^{3+} = 16.2 Ave TP = 6.5 Ave Al^{3+}/TP = 2.5	Al^{3+} = 1.5+2.2(TP) r = .553 n = 57 Ave Al^{3+} = 10.4 Ave TP = 4.0 Ave Al^{3+}/TP = 1.6	Insufficient data for regression n = 5 Ave Al^{3+} = 10.3 Ave TP = 6.2 Ave Al^{3+}/TP = 1.7	Al^{3+} = -.3+1.3(TP) r = .642 n = 15 Ave Al^{3+} = 7.2 Ave TP = 5.7 Ave Al^{3+}/TP = 1.3 Ave Residual TP = 1.2

* Iron was added as $FeCl_3$ except at two plants where pickle liquor was used.
** To achieve 1 mg/l TP.
TP = Average Total Phosphorus (mg/l as P) in the raw wastewater when chemical added to raw wastewater.
= Average Total Phosphorus (mg/l as P) in the secondary when chemical added to mixed liquor.
r = Correlation coefficient (the r's for all equations are significant at the 95% confidence level).
n = Number of observations.
** To achieve 1 mg/l TP.

Overall, the equations were capable of predicting chemical dosages for a particular wastewater to $\pm$ 30%. The implication of this is that in future treatability programs (Ontario or elsewhere) it may be possible to reduce the scope of the jar testing phase. For example, (Rush Engineering Services Limited 1979) in a recent treatability study at the Goderich, Ontario wastewater treatment plant utilized this approach to predict ferric chloride requirements.

Prested et al (1977) investigated the relationship between jar test predictions and full-scale performance for the Ontario data (Figure 6). Overall jar tests tended to over-estimate full-scale dosage requirements by about 15 to 25%.

Prested et al (1978) attempted to develop more sophisticated phosphorus removal predictive methodologies considering other wastewater characteristics such as hardness, alkalinity, conductivity, fluoride, and other factors. Even though the resultant predictive equations were somewhat better than the equations of Table 2 (i.e., $\pm$ 20% compared to $\pm$ 30%) the degree of improvement did not warrant the extra time and expense required for additional wastewater characterization.

UTILIZATION OF WASTE MATERIALS FOR PHOSPHORUS REMOVAL

Pickle liquors, waste aluminum chloride, carbide lime (by-product of acetylene production) and alumized red mud solids among other materials have been evaluated as phosphorus removal precipitants as part of the overall Great Lakes program. For details refer to reports by Fowlie and Shannon (1973) Shannon and Verghese (1975) and Scott (1973).

Many of the above mentioned waste materials were effective phosphorus precipitants, for example, Fowlie and Shannon (1973) concluded that pickle liquors, reclaimed pickle liquors ($FeSO_4.H_2O$) and carbide lime were capable of consistent phosphorus removal ($\leq$ 1/mg/L) at doses comparable to traditional iron and lime coagulants. However, other factors generally govern the use of waste materials including variability in composition of reaction chemical, content of impurities such as heavy metals which may have adverse side effects, and availability of material (quantity and geographic location).

For pickle liquors a prime concern is the build-up of heavy metals in the sludge streams. Fowlie and Shannon (1973) found that full-scale addition of pickle liquor at the Dunneville, Ontario treatment plant did not result in any increase in effluent heavy metals. However, waste sludge cadmium, chromium, copper, manganese, nickel, lead, zinc, arsenic and mercury levels all increased significantly from baseline (no pickle liquor addition). This factor could definitely influence the method of ultimate sludge disposal.

PHOSPHORUS REMOVAL FOR LAGOONS

Many smaller municipalities in Ontario and in Canada utilize lagoons (aerated, conventional stabilization and seasonal discharge) for treatment of their wastewater. Considerable effort during the Canada-Ontario program was devoted to developing appropriate phosphorus removal technologies to retrofit or supplement existing lagoon processes.

Continuous Over Flow Lagoon

For continuous discharge lagoons it was found alum or ferric chloride could be fed into the influent to the aerated cell. Frequently, this can be done at the raw sewage pumping station to ensure a high degree of mixing prior to reaching the lagoon. Again, the full scale design should be based upon a wide range of jar testing, excluding lime, carried out on the raw sewage as previously outlined. Lime has not been found to be effective (Graham and Hunsinger, 1977). Full-scale, temporary studies are generally not necessary. Several aerated lagoons in Ontario are now effectively achieving phosphorus removal. Some of the treatability results are summarized in Table 4.

TABLE 4

RESULTS OF PHOSPHORUS REMOVAL STUDIES AT LAGOONS IN ONTARIO

LOCATION	SIZE (ha)	CHEMICAL & DOSE	EFFLUENT QUALITY (mg/L)		
			BOD_5	TOTAL P	SUSPENDED SOLIDS
Sutton	6.5 Stabil. Pond	Ferric Chloride 20 mg/L as Fe	11	1	21
Shelburne	2.8 Stabil. Pond	Alum 225 mg/L	15-30	1	10 to 70
Waterford	6.1 Stabil. Pond	Lime 400 mg/L	15-40	1-3	10 to 90

Seasonal Retention Lagoons

The Ontario experience indicates that phosphorus removal can generally be achieved in seasonal retention lagoons by batch treatment of the lagoon contents with alum, or ferric chloride prior to discharge. Details of the technique are provided in a report by Graham and Hunsinger (1974).

The batch treatment dosages can be predicted with reasonable accuracy by jar testing, and the actual dosage selected for full-scale treatment should be that which is needed to achieve a jar test residual phosphorus of 0.5 mg/L or less.

In the batch treatment process, the chemical, as a liquid, is contained in a 6-700 litre plastic tank mounted amidships in a 70-HP outboard (or equivalent) boat. Chemical dispersal throughout the lagoon is by means of a 50 mm syphon discharging into the prop wash. The main application objectives in lagoon batch treatment are rapid, even distribution of the chemical and good mixing.

Batch chemical treatment of seasonal retention lagoons using alum and ferric chloride has been highly successful in Ontario. Effluent quality after treatment is usually less than 10 mg/L BOD and 20 mg/L SS with total phosphorus generally less than 0.5 mg/L. Due to a slight deterioration in effluent quality after about two weeks, it has been recommended to complete the discharge of the lagoon contents within eight to ten days after treatment.

Add-On Lagoon Treatment

The Environmental Protection Service conducted extensive phosphorus removal studies (Shannon and Salvo, 1974) at a stabilization pond in Quebec. A schematic of the pilot plant installation is shown in Figure 7.

The pilot plant results demonstrated that a chemical treatment system incorporating rapid mix, flocculation, tube settling and mixed-media filtration using alum as the prime coagulant was capable of producing a high quality effluent with less than 0.2 mg/L of total phosphorus. The stabilization pond effluent was subject to pronounced seasonal fluctuations (Figure 8) with winter turbidity, phosphorus and total nitrogen levels being 2 to 3 times higher than summer values. As a result it was recommended that the full-scale chemical conditions be adjusted as follows: alum dosages for summer and winter operation

should be 100 and 150 mg/L respectively. For pH control during winter a supplemental dose of soda ash Na_2CO_3 of 50 mg/L would be required.

A bench-scale evaluation of several settling and filter aid polymers revealed that activated silica was the most effective. Filter backwash requirements varied from 3 to 5% of the treated effluent volume and were found to be largely dependent on alum dosage.

It was demonstrated during this study that the problem of high effluent turbidities and associated aluminum losses to the effluent could be surmonted by proper selection of a polymer filter aid and the optimum selection of filter terminal headloss setting. (Note: The terminal headloss setting is that headloss at which backwashing will be initiated). Ideally the filter headloss and breakthrough turbidity should be attained simultaneously. Two typical turbidity breakthrough curves are shown in Figure 9. The curve for no polymer illustrates an undesirable pattern where the filter run is short and there is a rapid rise in turbidity shortly after the filter starts up. Although the pattern with polymer is not ideal, it demonstrates the effect of activated silica on increasing filter run time.

HIGH LIME TREATMENT

It is apparent from the previous discussion that most chemical phosphorus removal systems use either iron salts or alum as the precipitant. In the early stages of the Ontario treatability program, it appeared that lime would have a significant share of the total number of phosphorus removal installations, since it was generally the lowest cost precipitant. However, throughout the treatability program it became apparent that very few if any full-scale installations would use lime.

The main factors against using lime appear to have been (1) operators find it more difficult to handle than alum or ferric chloride, i.e., slurry versus liquid, (2) designers have had difficulty in developing a low maintenance lime system for wastewater treatment and (3) only the "low lime" process (pH 9.5 - 10.0) is directly compatible with conventional secondary treatment plants, a high lime effluent would require neutralization (recarbonation) prior to biological treatment.

In spite of these drawbacks the high-lime process has demonstrated that it has a high wastewater treatment potential. In this regard pilot plant studies were conducted at the Wastewater Technology Centre in Burlington comparing the treatment of a municipal wastewater with a high-lime process versus conventional biological treatment (extended aeration) with phosphorus removal. The results have been presented in a report by Brouzes et al (1974).

The average effluent quality for the two systems is summarized in Table 5.

TABLE 5

COMPARISON OF EFFLUENT QUALITY OF A HIGH LIME PROCESS VERSUS AN EXTENDED AERATION SYSTEM WITH P-REMOVAL

	BOD_5		TOTAL-P		SUSPENDED SOLIDS	
	mg/L	% Red.	mg/L	% Red.	mg/L	% Red.
Raw Wastewater	95		4.7		159	
High Lime[1]	20	79	0.3	93	19	88
Extended Aeration	5	95	0.4	91	5	97

[1]After carbonation of final effluent.

Although the high lime system did not achieve the same degree of BOD_5 and suspended solids removal as the extended aeration plant it still produced an effluent with consistently low BOD_5, suspended solids and phosphorus. See Figures 10 and 11 for respective time series plots of BOD_5 and total phosphorus for the two plants.

In terms of heavy metals removal and viral and bacterial inactivation the high lime process produced an effluent which was superior to the biological system. A comparison of the bacteriological quality of the two effluents is shown in Table 6.

The above data point out that the high lime process could possibly find application for upgrading or retrofiting existing primary systems where the receiving water does not require a high degree of BOD_5 and SS removal but high phosphorus removals are required.

TABLE 6

BACTERIAL COUNTS IN HIGH LIME PROCESS
AND EXTENDED AERATION PLANT EFFLUENTS

	TOTAL COLIFORMS (#/100 ml)	FECAL COLIFORMS (#/100 ml)	FECAL STREPTOCOCCI (#/100 ml)
Raw Wastewater	2.7×10^6	9.8×10^5	1.5×10^5
High Lime	3.8×10^2	2×10^1	3×10^3
Extended Aeration	5.1×10^4	6.6×10^3	3.0×10^4

REFERENCES

Boyko, B.I., and J.W.G. Rupke, "Phosphorus Removal Within Existing Wastewater Treatment Facilities", Canada-Ontario Agreement Research Report No.44, (1976)

Brouzes, R.J.P., R.J. Neufeld and B.P. LeClair, "The Effect of High Lime Treatment on Raw Sewage With Special Reference to the Destruction of Bacteria and Survival of Fish", Unpublished paper, Domtar Research Centre, Senneville, P.Q. (1974)

Fowlie, P.J.A., and E.E. Shannon, "Utilization of Industrial Wastes and Waste By-products for Phosphorus Removal and Inventory and Assessment", Canada-Ontario Agreement Research Report No. 6, (1973)

Graham, H.J. and R.B. Hunsinger, "Phosphorus Removal in Seasonal Retention Lagoons by Batch Chemical Precipitation", Canada-Ontario Agreement Research Report No. 13, (1974)

Graham, H.J. and R.B. Hunsinger, "Phosphorus Reduction from Continuous Overflow Lagoons by Addition of Coagulants to Influent Sewage", Canada-Ontario Agreement Research Report No. 65, (1977)

Ministry of the Environment (MOE), "Guidelines for Conducting Treatability Studies for Phosphorus Removal at Wastewater Treatment Plants", April, 1972. (Printed in COA Conf. PRoc. No. 1, Phosphorus Design Seminar Toronto, Ontario May 28, 29, 1973)

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

Prested, B.P., E.E. Shannon and R.J. Rush, "Development of Prediction Models for Chemical Phosphorus Removal, Volume II, Canada-Ontario Agreement Research Report No. 78, (1978)

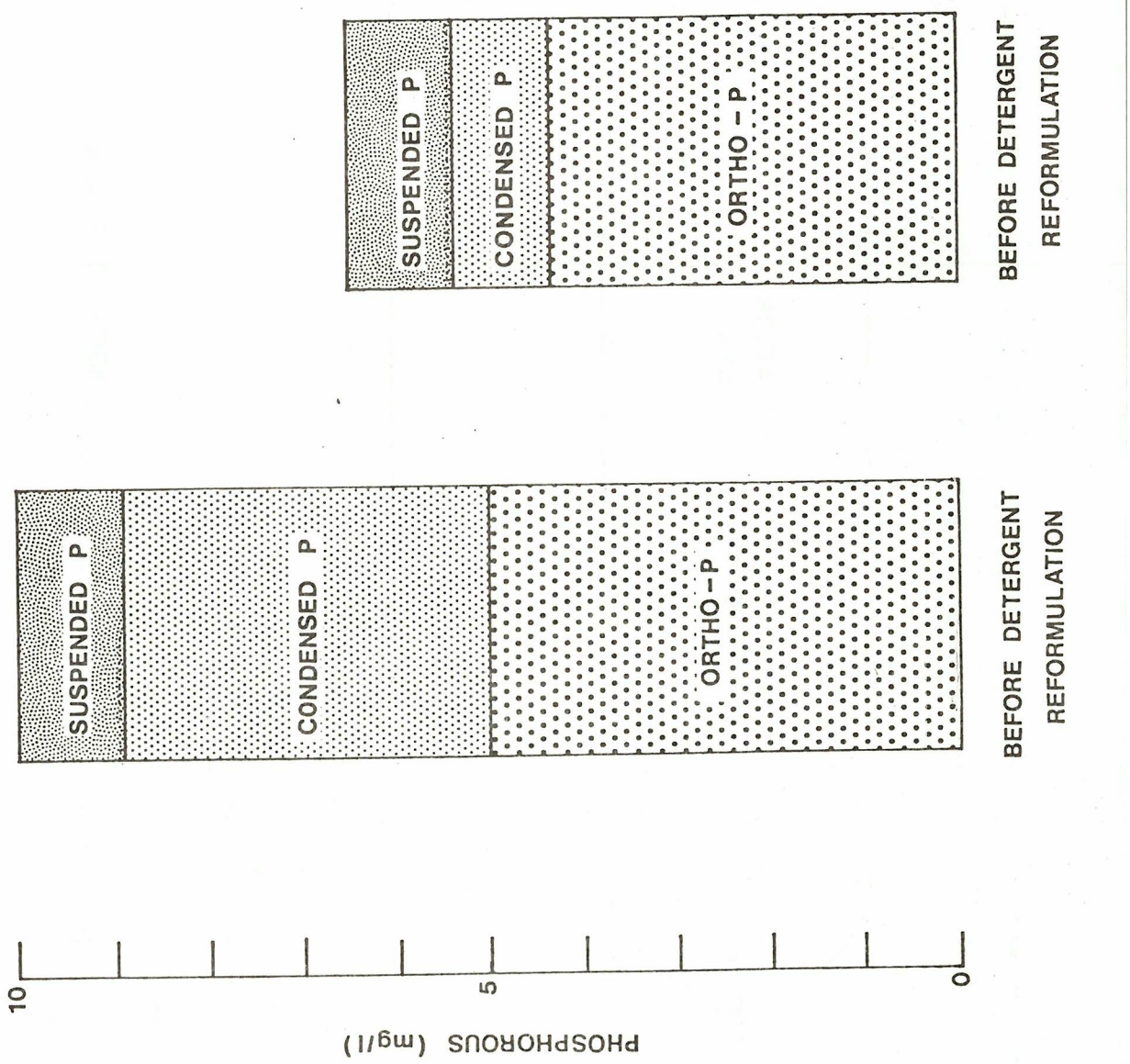
Rush Engineering Services Limited, "Phosphorus Removal Considerations for Expansion of the Goderich Wastewater Treatment Plant", Unpublished report (1979)

Scott, D.S., "Use and Production of Iron Salts for Phosphorus Removal", Canada-Ontario Agreement Research Report No. 5, (1973)

Schmidtke, N.W. and D.B. Cohen, "Municipal Sludge Disposal on Land, A Down-to-Earth Solution", Presented at the 29th Annual Convention of Western Canada Water and Sewage Conference, Edmonton, Alberta, Sept. 29 and 30, 1977.

Shannon, E.E. and J.M. Salvo, "Tertiary Phosphorus Removal and Limiting Nutrient Studies at C.F.S. Lac Ste. Denis", Environmental Protection Service Report EPS4-WP-74-1, (1974)

FIGURE 1. FORMS OF PHOSPHORUS IN DOMESTIC WASTEWATER



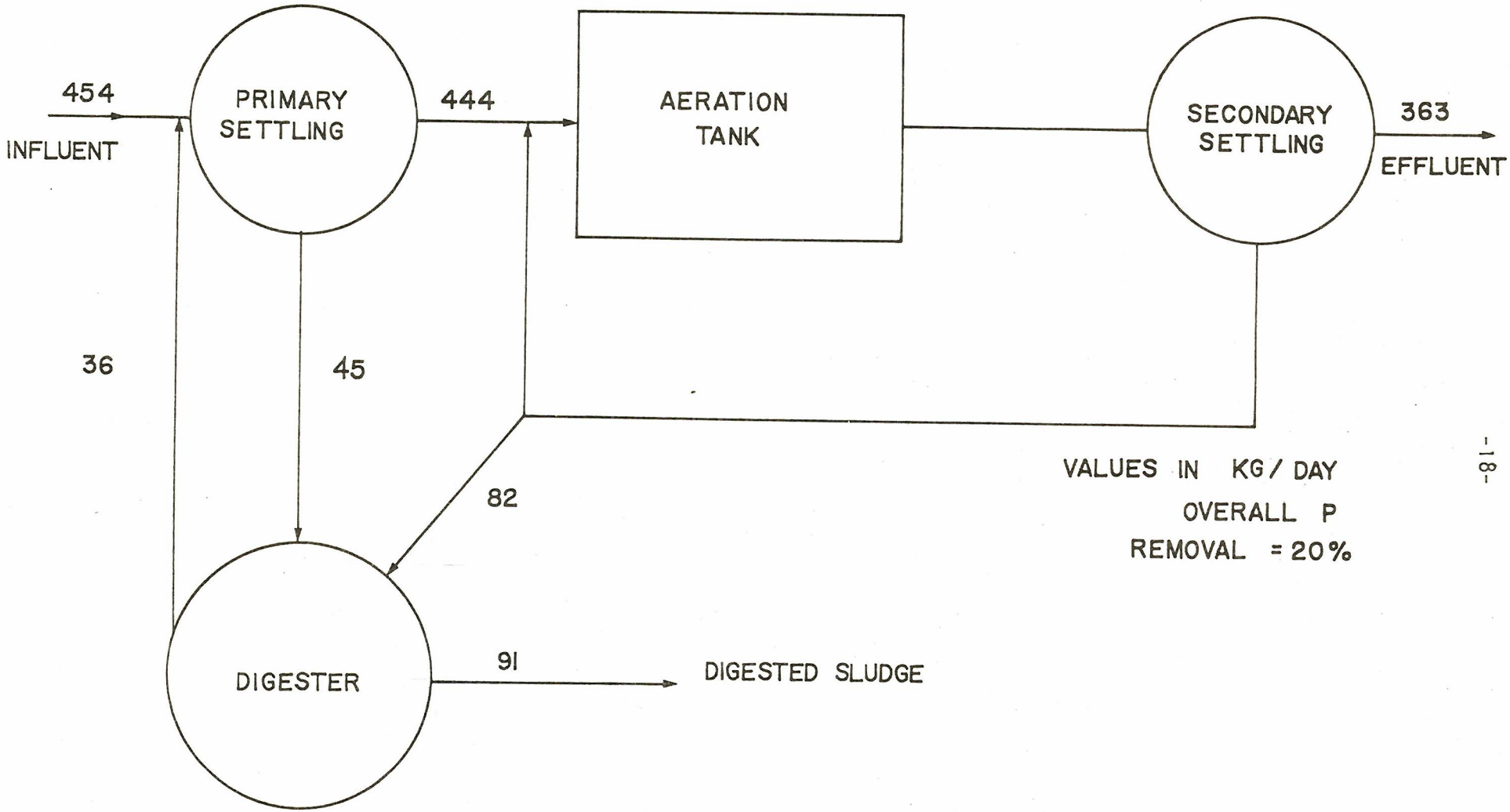
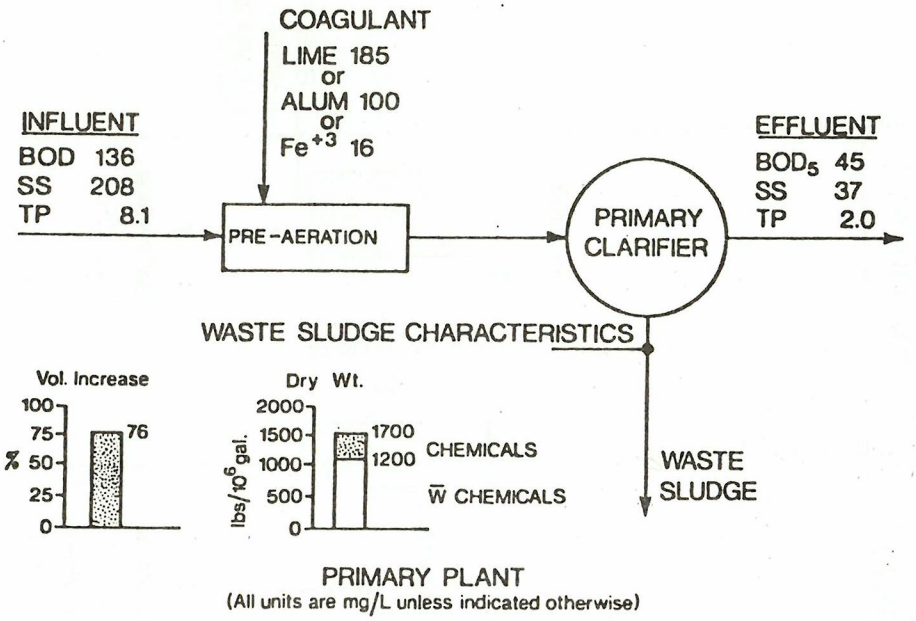


FIGURE 2. FATE OF PHOSPHORUS DURING CONVENTIONAL TREATMENT



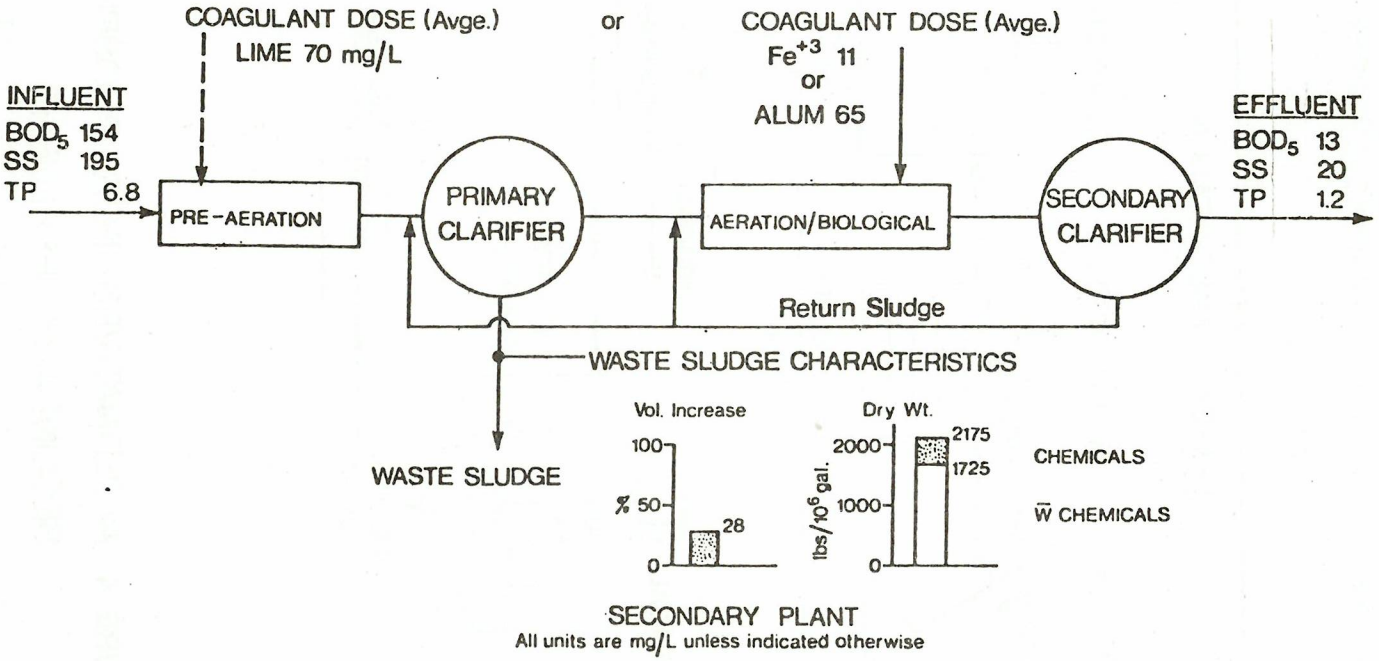


FIGURE 3. TYPICAL PHOSPHORUS REMOVAL PROCESSES
(after Schmidtke and Cohen, 1977)

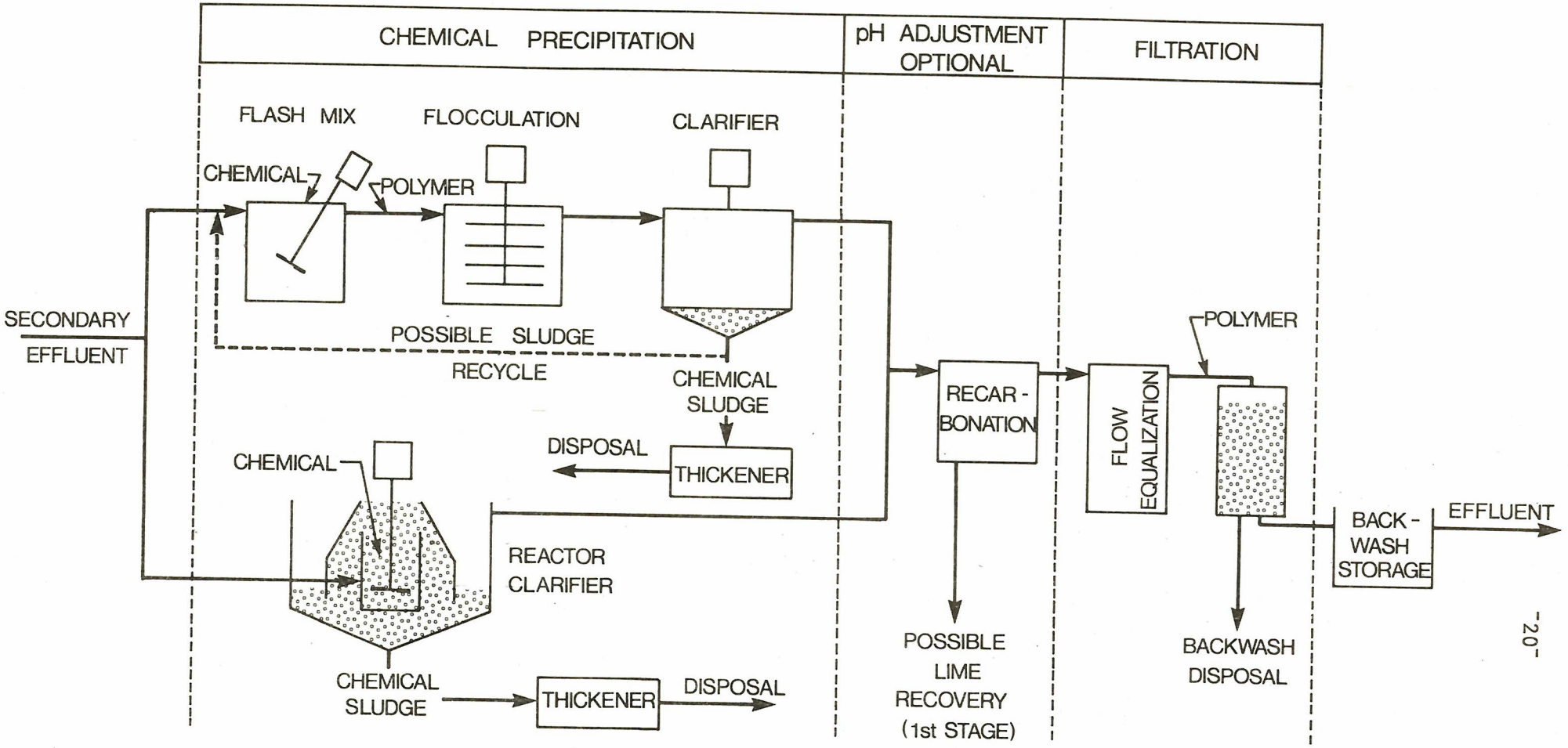


FIGURE 4. ALTERNATIVES FOR PHOSPHORUS REMOVAL FROM A SECONDARY EFFLUENT

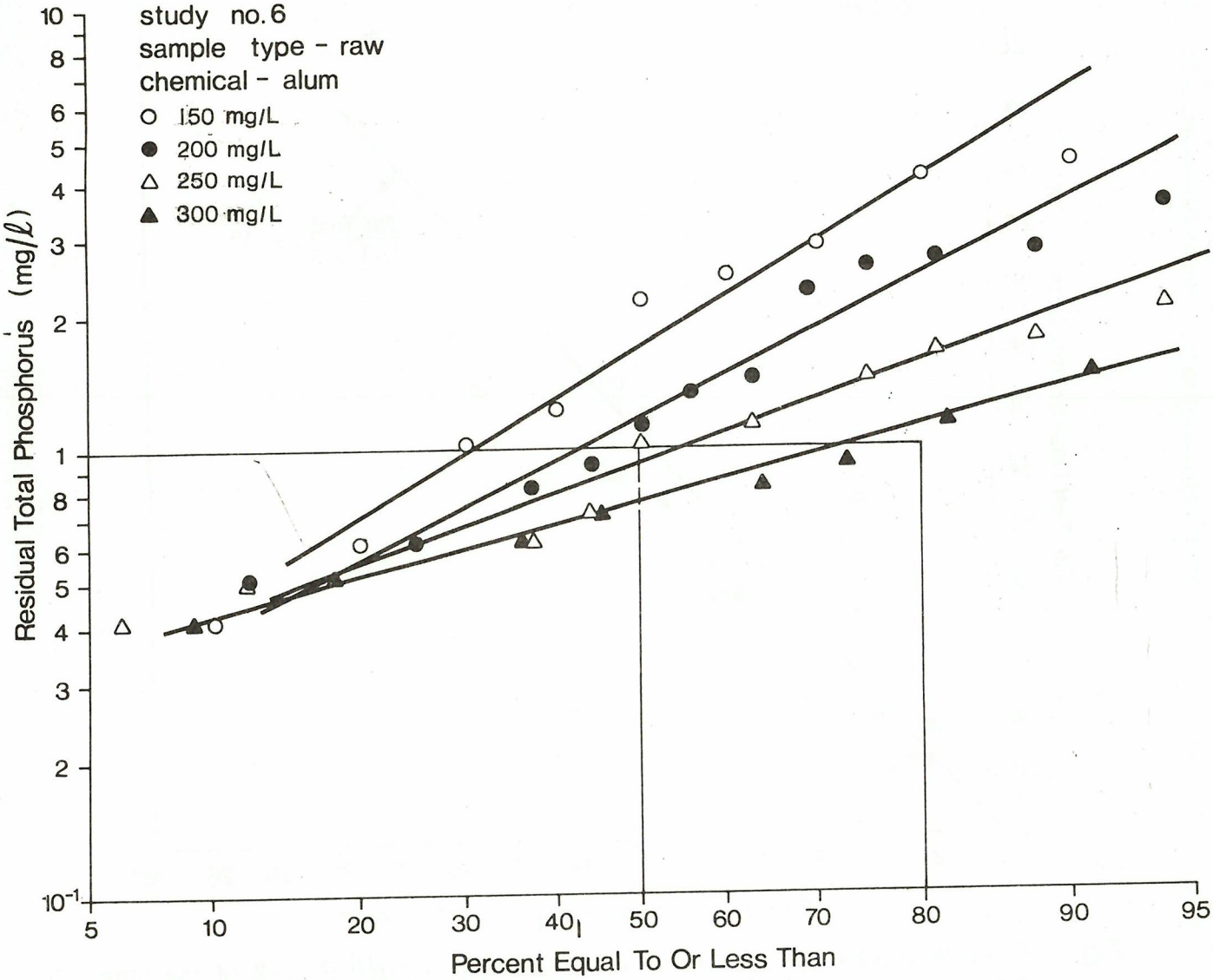


Fig.5. PROBABILITY PLOT FOR KITCHENER JAR TEST TREATABILITY DATA

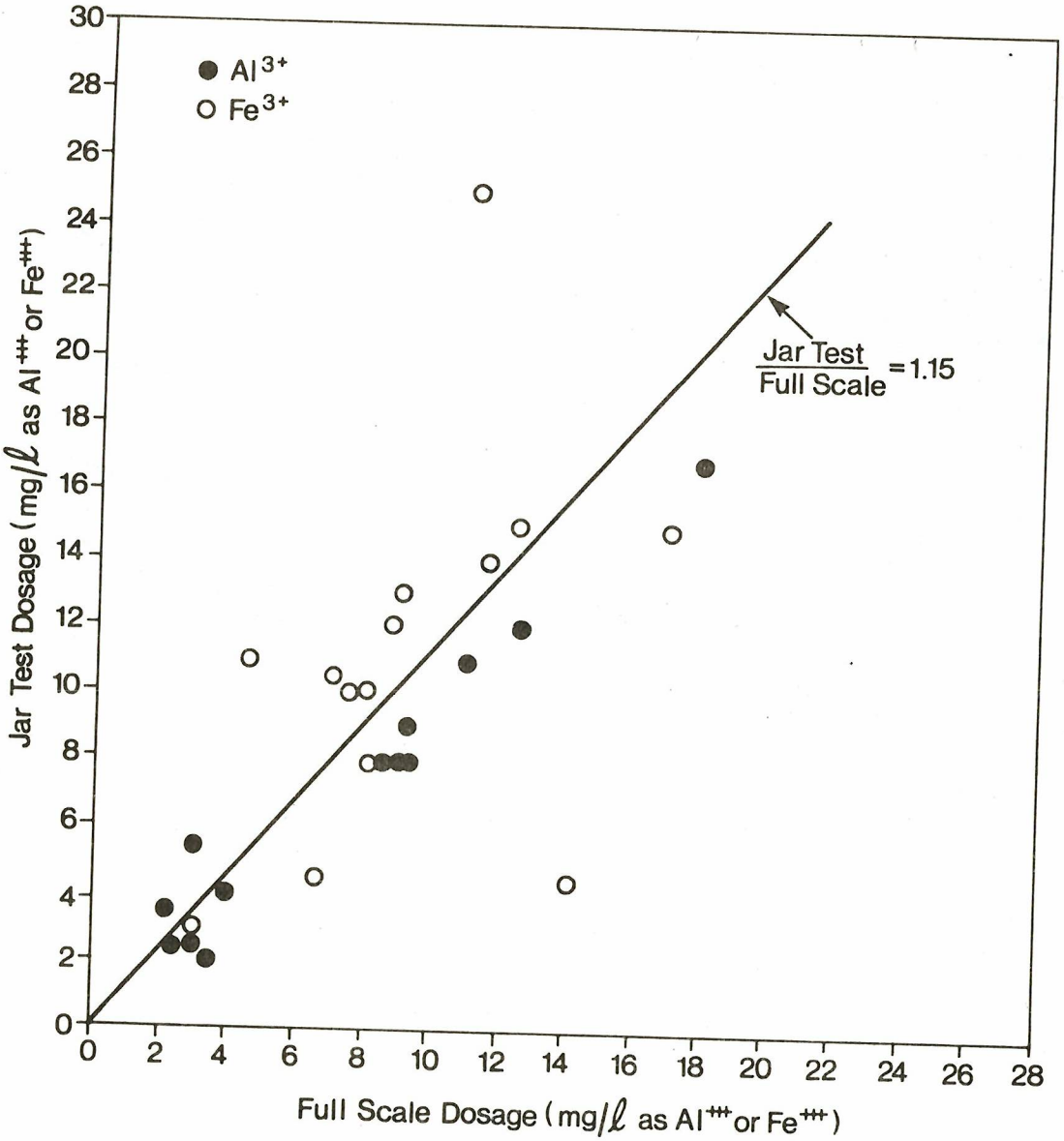


Fig.6. COMPARISON OF JAR TEST DOSAGES (for 1 mg/ℓ P 50 % of the time) TO FULL SCALE DOSAGES (for 1 mg/ℓ P average) AT PLANTS ADDING ALUM OR FERRIC CHLORIDE TO THE MIXED LIQUOR.

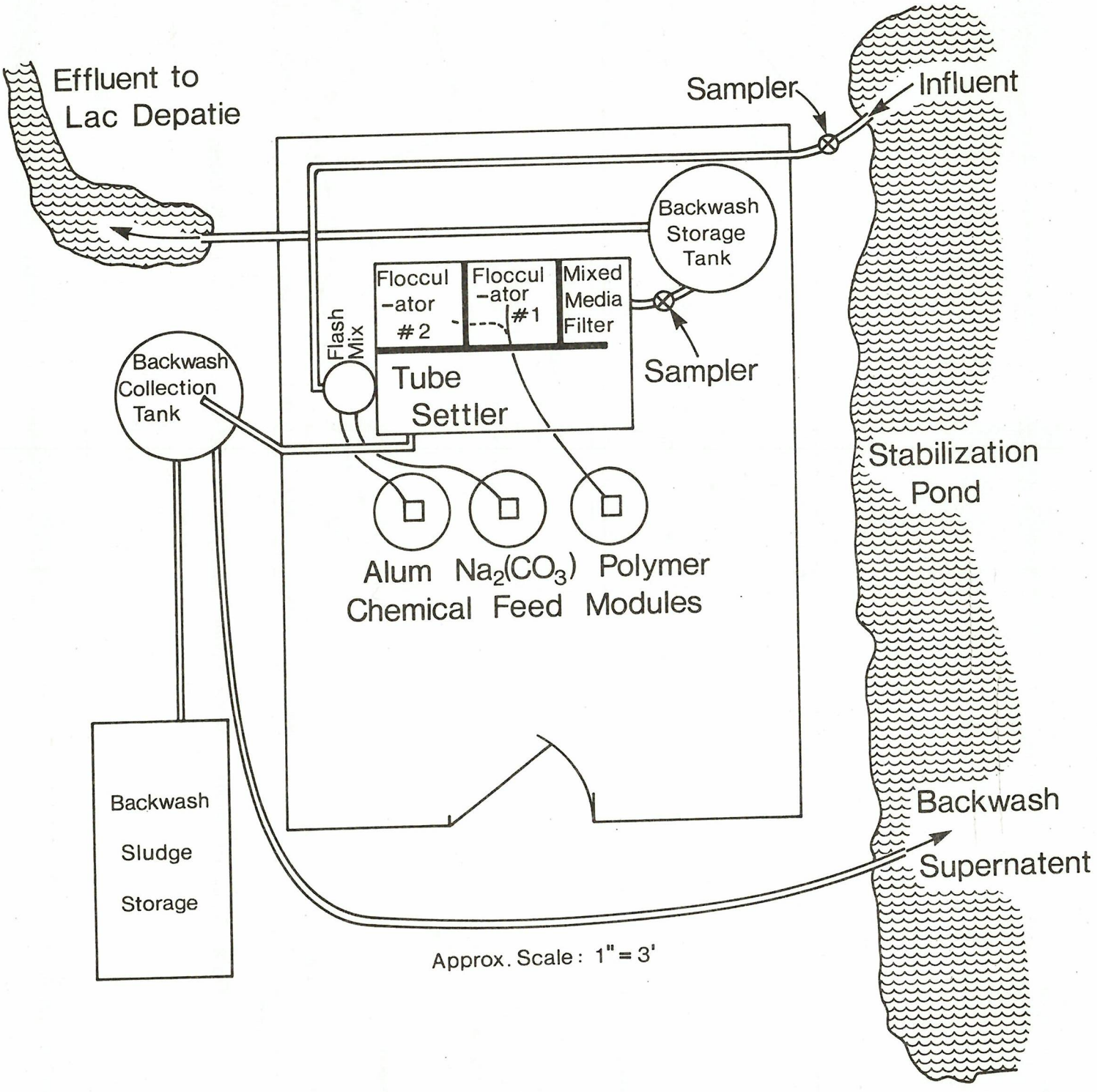


Fig. 7. Schematic of pilot plant installation at Lac St. Denis

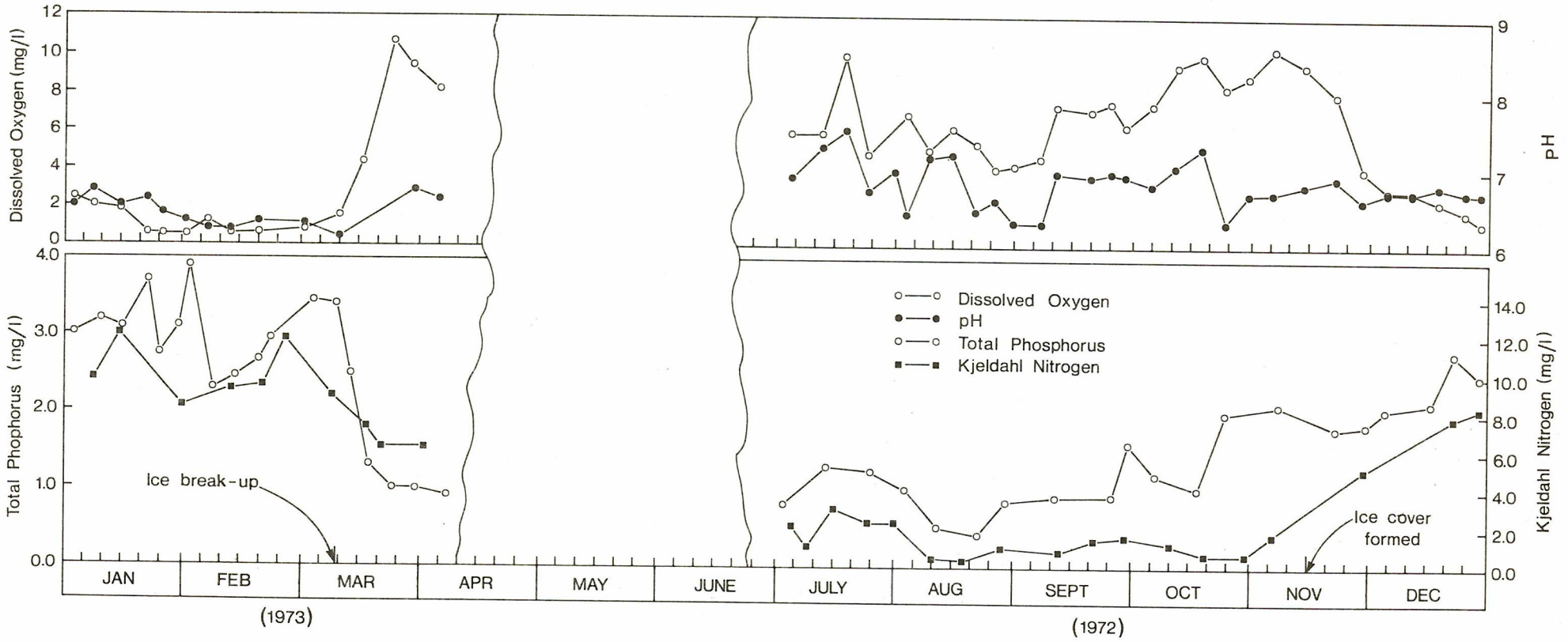


Fig. 8. Seasonal variations in the effluent quality of the stabilization pond at Lac St. Denis

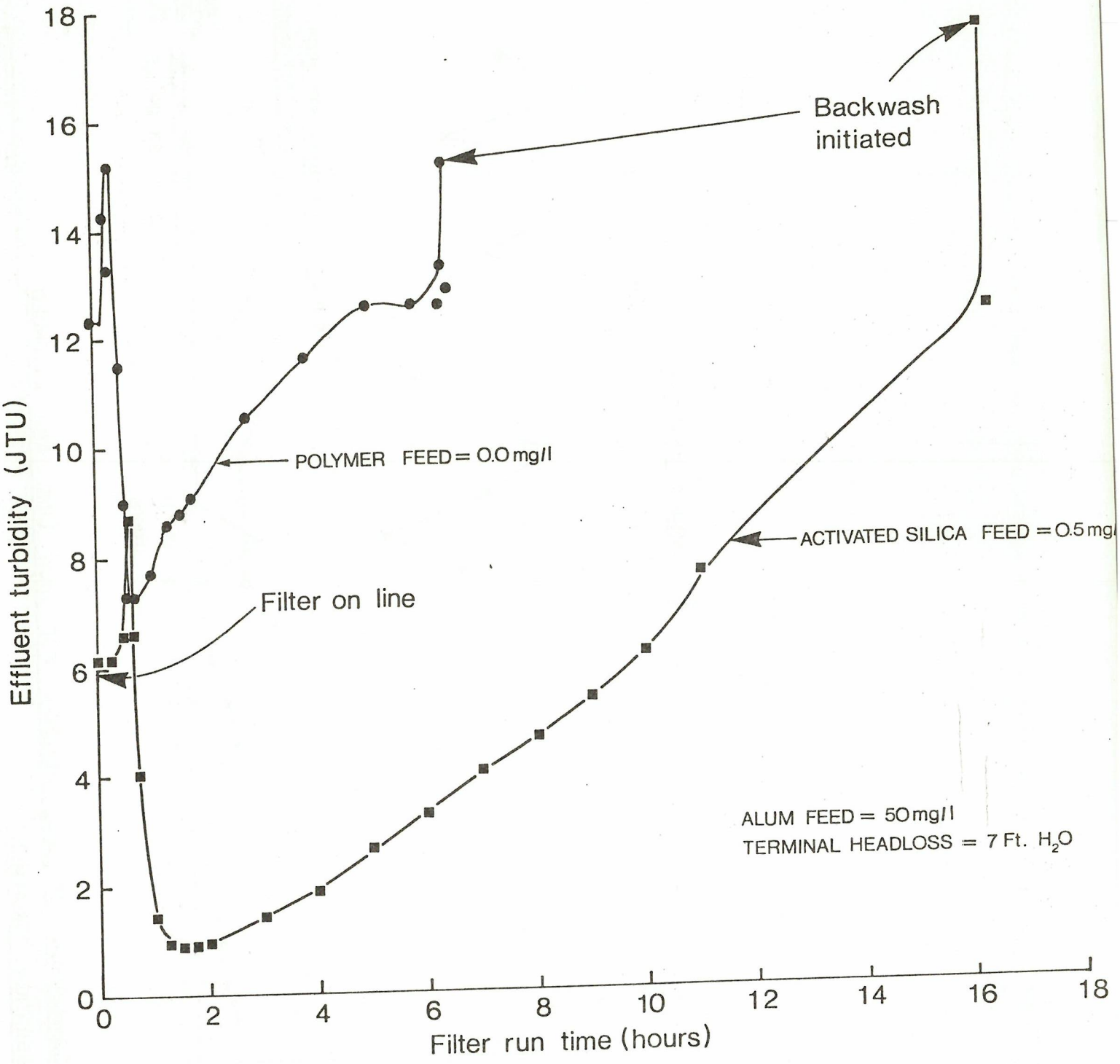


Fig. 9. Turbidity breakthrough on mixed media filter

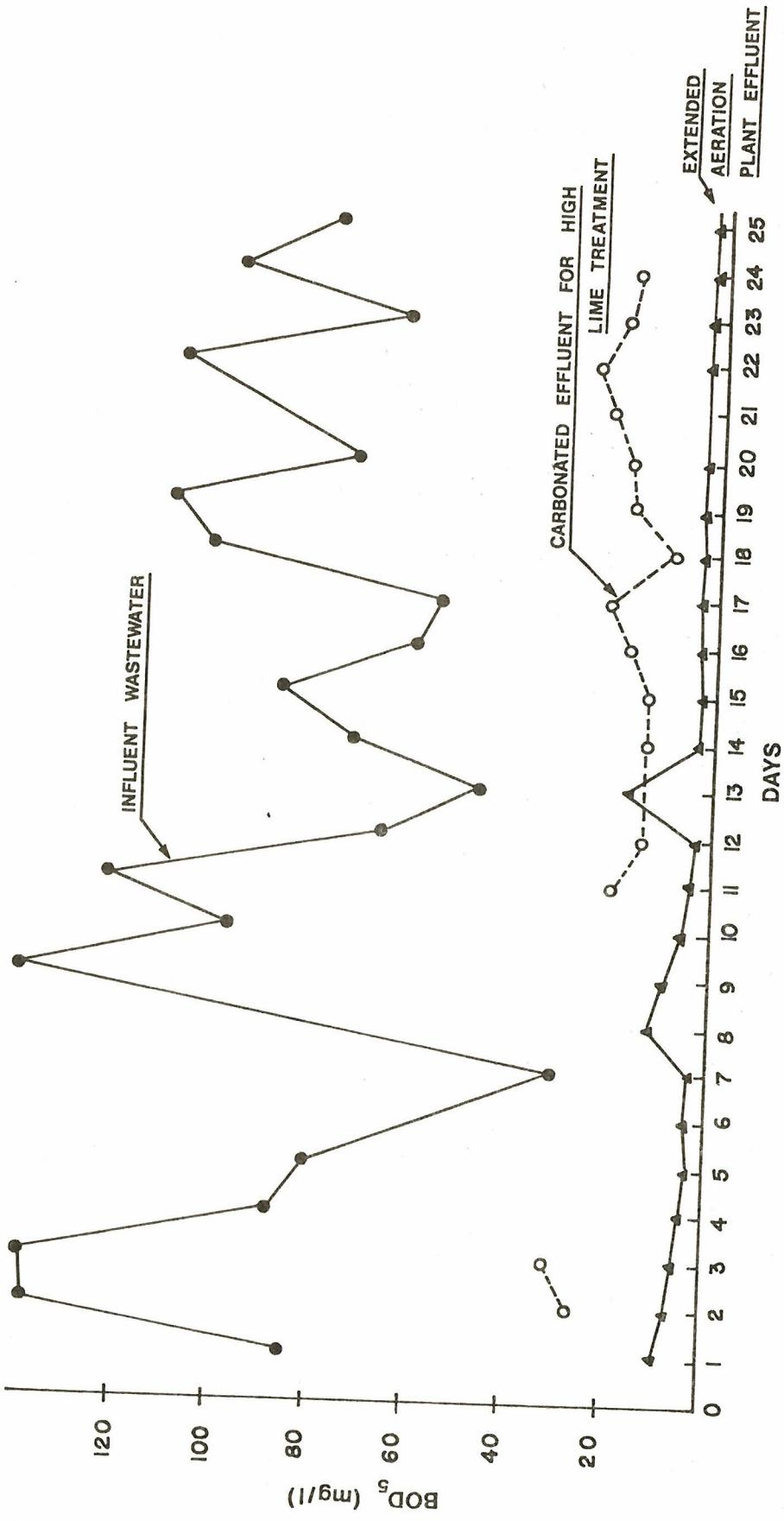


FIG. 10. COMPARISON OF BOD_5 REMOVAL FOR HIGH LIME AND EXTENDED AERATION TREATMENT

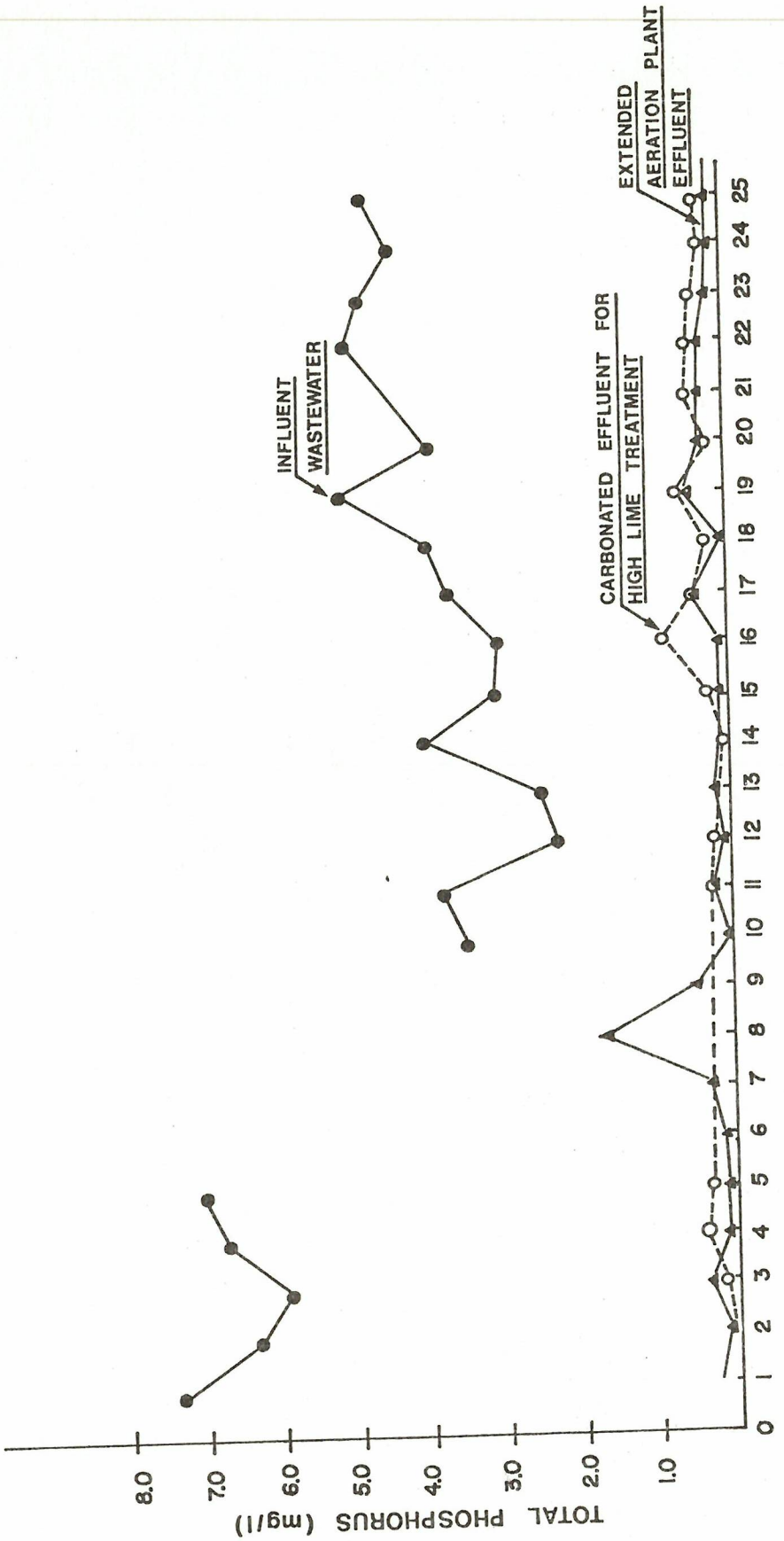


FIG. 11. COMPARISON OF TOTAL PHOSPHORUS REMOVAL FOR HIGH LIME AND EXTENDED AERATION TREATMENT

factors affecting phos. removal

(1) Chemical - alkalinity, hardness, ionic strength

(2) Physical - temp.

(3) hydraulic - mixing, flocculation, overflow rate, T.C.

NUTRIENT REMOVAL IN SUSPENDED GROWTH SYSTEMS WITHOUT CHEMICAL ADDITION - CONTROLLING FACTORS AND EFFECTS ON DESIGN

by

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ABSTRACT

Two pilot scale single sludge nitrification denitrification sytems incorporating provisions expected to promote a high degree of phosphorus removal without chemical addition were operated in parallel with a nitrification system functioning as a control unit. The study was aimed at obtaining quantitative information on how various factors affect the degree of phosphorus removal and determining the effect of anoxic and anaerobic conditions on the growth rates of nitrifiers. The results indicated that the anoxic and anaerobic conditions have no effect on nitrification and that a high degree of phosphorus removal can be obtained which is dependent on the degree of COD removal.

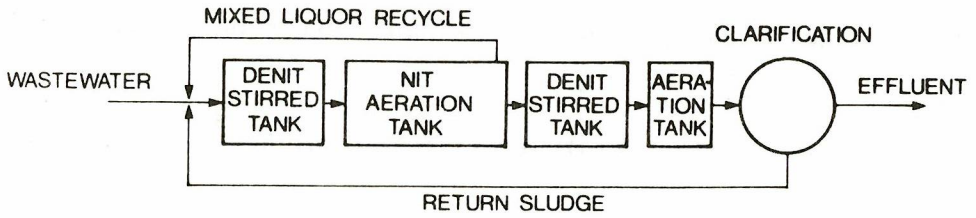
NUTRIENT REMOVAL IN SUSPENDED GROWTH SYSTEMS WITHOUT CHEMICAL ADDITION - CONTROLLING FACTORS AND EFFECTS ON DESIGN

Bruce E. Jank and Derek Vachon, Wastewater Technology Centre, Burlington, Ontario.

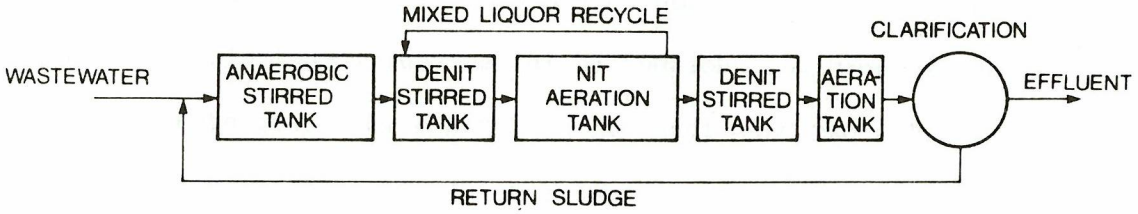
Paul M. Sutton, Dorr-Oliver, Inc., Stamford, Connecticut.

INTRODUCTION

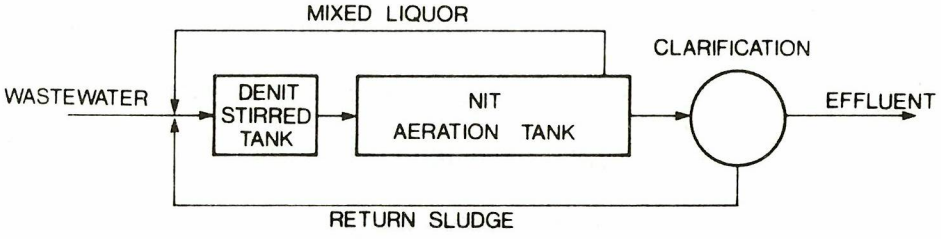
The need for nutrient control in the treatment of municipal and industrial wastewater has been well documented in numerous technical reports and papers[1,2]. Recent attention has been focused on single sludge suspended growth systems which combine the processes of carbon oxidation, nitrification, and denitrification in one or more reactors without intermediate clarification and without the addition of an external carbon source for denitrification. It has been reported that in certain single sludge configurations designed specifically for nitrogen removal, in excess of 90 percent phosphorus removal can be obtained without chemical addition[3]. Pre-denitrification or nitrate removal at the head end of the process using primary or raw wastewater as the carbon and energy source is the principal mechanism for nitrogen removal in all these systems. In the configuration referred to as the Bardenpho process[3] (Figure 1, configuration 1), the mixed liquor is exposed to an anaerobic (absence of oxygen and nitrate) time period followed by an aerobic period. The anaerobic phase is preceded by an anoxic (absence of oxygen) period during which nitrate reduction or denitrification occurs. During the anaerobic-aerobic sequence, biological P release and uptake occurs. The net P removal is reported to be in excess of that normally observed in activated sludge systems. On the premise that for enhanced P uptake the mixed liquor must go through an anaerobic-aerobic sequence and that anoxic conditions existing during de-



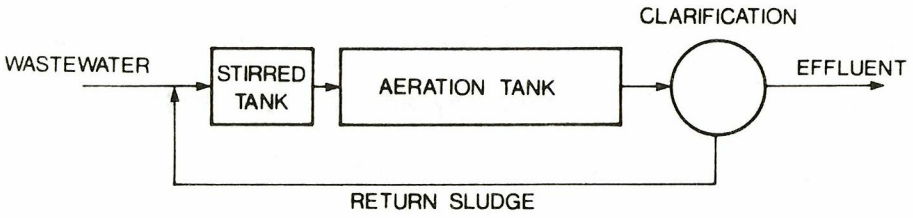
CONFIGURATION 1. BARDENPHO N AND P REMOVAL PROCESS[3]



CONFIGURATION 2. "MODIFIED" BARDENPHO PROCESS[9]



CONFIGURATION 3. PRE-DENITRIFICATION AND P REMOVAL SYSTEM[3]



CONFIGURATION 4. INCORPORATION OF P REMOVAL IN NON-NITRIFYING SYSTEM[3]

Note, NIT refers to nitrification and DENIT to denitrification

FIGURE 1. SINGLE SLUDGE SYSTEMS INCORPORATING PHOSPHORUS REMOVAL

nitrification are detrimental to P release, variations of the original configuration have been proposed which may or may not include nitrogen removal (Figure 1).

The principle of promoting phosphorus release by exposing the microbial population to anaerobic conditions has been applied in the Phostrip process[4]. In this system a separate phosphorus stripping reactor is necessary and sludge retention periods in excess of 10 hours[5] are normally required to obtain high degrees of P release. The P in the supernatant from the stripper is then precipitated by chemical addition. In a recent study of the Phostrip process[6], it was shown that the stripping tank played the critical role in assuring high levels of P removal and not luxury uptake of phosphorus into the sludge. The phosphorus concentration of the sludge volatile suspended solids in the study was approximately two percent. In a full scale study by Levin and his co-workers[7], total phosphorus was reduced from influent values of greater than 5 mg/ℓ P to effluent values of less than 1 mg/ℓ using the Phostrip concept. In the study the anaerobic P stripping reactor was operated at a hydraulic retention time of approximately 18 hours. The P concentration of the sludge volatile suspended solids was approximately three percent. This is the P content that the design of the Phostrip process is based on[8].

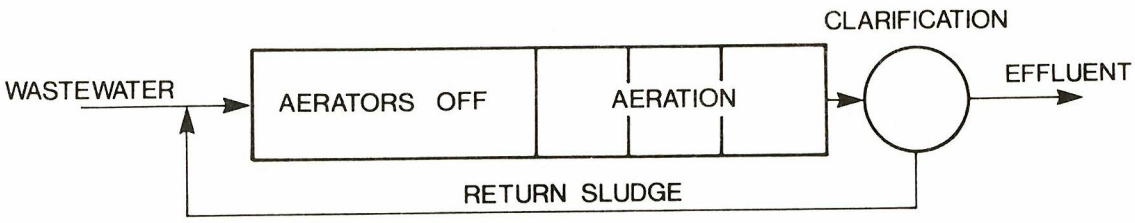
Although it has been shown that successful operation of the Phostrip process essentially relies on operation of an anaerobic stripping tank at long retention periods followed by phosphorus removal by chemical precipitation, the Bardenpho and other configurations (Figure 1) strictly rely on accumulation of phosphorus in the biomass for removal. In developing the Bardenpho process, Barnard achieved effluent filtered phosphorus levels below 1 mg/ℓ.[3] In his studies the total actual (based on influent plus re-

cycle flow rates) anoxic plus anaerobic hydraulic retention time (HRT) was 2.33 hours. This corresponded to a nominal (based on influent flow rate) HRT of six hours. Because the anaerobic conditions required for P release were achieved in the reactors in which initially nitrate reduction (anoxic conditions) occurred, it is not possible to determine the actual anaerobic conditions. The total actual and nominal aerobic HRT's were 1.5 and 7 hours, respectively[3]. Although claiming that P removal was due to luxury uptake no measurements were made on the P content of the sludge.

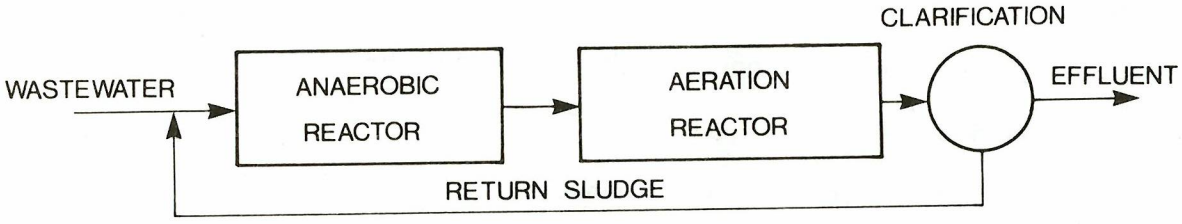
McLaren and Wood[9] investigated the Bardenpho process configuration at the pilot scale level and did not obtain the high phosphorus removal previously reported. Effluent total P levels of approximately 4 mg/ℓ or 50 percent P removal were observed. However, by incorporating an anaerobic stage preceding the first stage reactor of the Bardenpho process (Figure 1, configuration 2) P release and subsequent luxury uptake was observed resulting in effluent total P levels below 1 mg/ℓ. The nominal HRT in the first stage anaerobic reactor of the configuration initially was four hours (actual HRT two hours). The authors claim that P release and subsequent luxury uptake was still maintained when the nominal HRT in the first stage reactor was reduced to one hour (actual HRT 0.5 hours) and the second stage anoxic reactor was eliminated. They concluded that the P removal mechanism was purely biological relying on the development of the "proper" organism population. No values were presented for the P content of the biological sludge. The study of McLaren and Wood was followed by a large pilot scale by Simpkins and McLaren[10]. Operating the pilot plant in the "modified" Bardenpho configuration they were able to obtain effluent orthophosphate (PO_4-P) levels of approximately 1 mg/ℓ "most of the time". The average

influent total phosphate as P was approximately 5 mg/ℓ. The first stage anaerobic reactor in their study was operated at a nominal HRT of 2.4 hours or actual HRT of 1.2 hours. In discussing their results they concluded that bacterial uptake and not chemical precipitation was the primary mechanism accounting for the enhanced P removal. Although no values for the P content of the biological sludge were presented it can be calculated that in order to obtain the P removals reported, the level must be approximately 7 to 8 percent (P/VSS x 100).

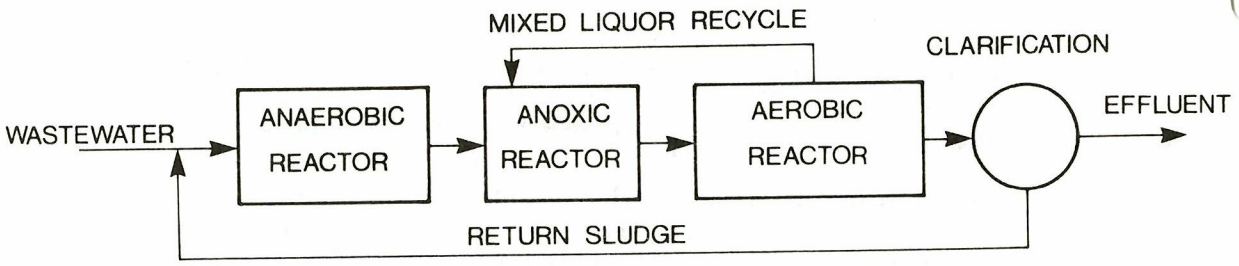
In recent years the city of Johannesburg, South Africa, has been involved in extensive pilot and full scale studies[11,12,13] designed to determine the degree of nutrient removal which can be achieved in the modified Bardenpho process and in other single sludge configurations (Figure 2). Each configuration incorporates, in principle, a first stage anaerobic step during which P release from the sludge occurs followed by a second stage aerobic step for P uptake. The mechanism responsible for enhanced P uptake was related to development of a selective organism population. The Johannesburg researchers have qualitatively identified factors which they claim will promote the development and maintenance of this population. In addition to the basic concept of creating an anaerobic step, of an actual HRT of 0.5 to 2 hours[11], followed by an aerobic step, they claim that a major factor contributing to enhanced biological P removal is the COD concentration fed to the system[13]; an increase in the COD concentration resulted in an increase in phosphorus removal. Their results[13] indicate that the aerobic hydraulic retention time is not critical as enhanced P uptake was observed in both nitrifying extended aeration and high rate activated sludge plants in which nitrification did not occur.



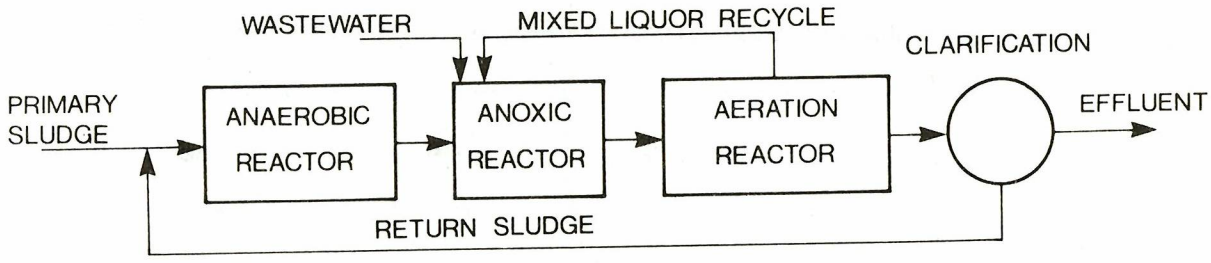
MODIFIED EXTENDED AERATION FULL SCALE ACTIVATED SLUDGE PLANT FOR N AND P REMOVAL[12]



MODIFIED HIGH RATE ACTIVATED SLUDGE PILOT PLANT FOR P REMOVAL[13]



MODIFIED PRE-DENITRIFICATION PILOT PLANT FOR N AND P REMOVAL[13]



PRIMARY SLUDGE ADDITION TO MODIFIED PRE-DENITRIFICATION PILOT PLANT FOR N AND P REMOVAL[13]

FIGURE 2: SINGLE SLUDGE REMOVAL CONFIGURATIONS INVESTIGATED BY JOHANNESBURG RESEARCHERS

Extensive laboratory investigations into nitrogen and phosphorus removal in single sludge systems have been carried out in recent years at the University of Cape Town in South Africa, by Marais and numerous co-workers[14, 15, 16]. Martin and Marais[14], operating laboratory reactors in the Bardenpho configuration, concluded that the first stage anoxic reactor played the critical role in ensuring enhanced P removal. In contrast to the studies of Barnard[3] and other South African researchers[11, 13], Martin and Marais found that P removal occurred in the first stage anoxic reactor and did not rely on P release and subsequent uptake in the aerobic reactor. In this regard they concluded that the presence of nitrate in the anoxic reactor had no influence on P removal. They did agree that an increase in the influent COD concentration will result in an increase in P removal. In their study the maximum removal of P was obtained at a first stage actual HRT of approximately 30 minutes. Marsden and Marais[16] conducted investigations aimed at specifically determining the role played by the first stage anoxic reactor in enhanced P removal. Studying the pre-denitrification nitrification configuration (Figure 1, configuration 3) the authors confirmed Martin and Marais' findings that P removal occurred in the first stage anoxic reactor and that optimal removal occurred at an actual HRT of 30 minutes. Expressing their results in terms of the ratio $\Delta P/\Delta COD$ based on influent and final effluent values, they found that the ratio was "far in excess of normal sludge requirements". Furthermore, they concluded that the system solids retention time (SRT) had an effect on this ratio. When the system SRT was reduced from 20 days to 10 days the ratio of $\Delta P/\Delta COD$ increased. The system SRT was defined as the total volatile suspended solids in the reactors divided by the daily volatile suspended solids lost in the effluent or intentionally wasted.

The efficiency of biological nitrogen removal is correlated to the degree of nitrification achieved. The autotrophic nitrifiers have lower yields and growth rates, and are more sensitive to environmental conditions than the heterotrophic denitrifiers. Consequently, nitrification probably will prove to be the crucial step in any single sludge nitrification and denitrification process. Incorporating P removal into single sludge systems by exposing the mixed liquor to a period of anaerobic conditions in addition to the anoxic conditions required for nitrate removal may adversely affect the nitrifier growth rate. Marsden and Marais[16] previous to studying the pre-denitrification nitrification configuration operated laboratory reactors in the Bardenpho configuration (Figure 1, configuration 1) primarily to study nitrate removal in the anoxic reactors. At a total anoxic HRT (actual) of 1.7 hours, nitrification was severely inhibited even though the aerobic reactor SRT was 10 days and the temperature 14°C. The aerobic SRT was determined from the fraction of suspended solids under aerobic conditions multiplied by the system SRT. According to the U.S.-EPA[1] the minimum sludge age for nitrification at 14°C is approximately six days. The results of Marsden and Marais[16] are in contrast to the results of Wuhrmann[18] who claimed that anoxic reactor retention times for denitrification of up to five hours had no effect on the nitrification process. It is difficult to assess the effect on nitrification of anaerobic conditions from the literature as in most cases either the SRT and temperature conditions are not reported or the aerobic SRT of the system is at least two to three times the minimum required for nitrification[1] at the operating temperature.

In view of the conflicting reports on the effects of various factors on nutrient removal in single sludge systems, a pilot plant study was con-

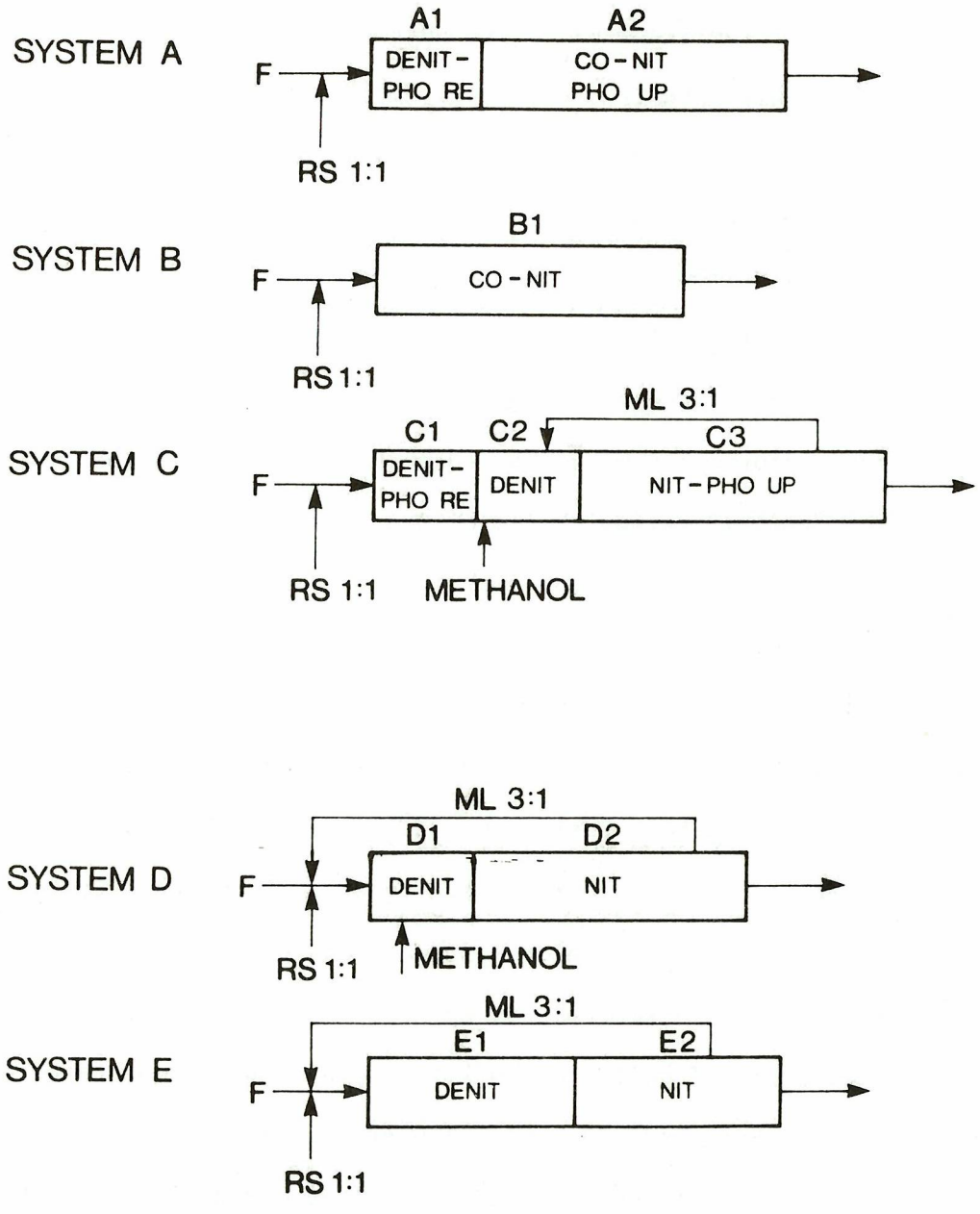
ducted at the Wastewater Technology Centre, Environmental Protection Service, Burlington, Ontario. The study was aimed specifically at;

1. obtaining quantitative information on how various factors such as process configuration, influent waste characteristics, temperature, system SRT, and mixed liquor retention period, affect the degree of P removal in single sludge systems, and
2. determining the effect of anoxic and anaerobic conditions on the growth rates of the nitrifiers.

In addition, information was obtained on N removal in single sludge systems in which denitrification relied on the organics present in the wastewater and/or an external carbon source.

EQUIPMENT AND PROCEDURES

Three parallel single sludge systems (Figure 3, systems A, B, and C) were operated according to a pre-determined experimental design over a period of approximately four months. Systems A and C were designed to achieve nitrification, partial denitrification, and enhanced P uptake. System B was designed to achieve nitrification alone and was operated as a control. Additional information on P removal was derived from systems D and E (Figure 3) which were designed specifically to study N removal and were operated during a previous study[19]. Systems C, D, and E achieved a high degree of N removal by returning sludge and a large volume of nitrified mixed liquor (Figure 3) to the denitrification reactors. In systems C and D the rate of denitrification was maximized by ensuring that the reaction proceeded under non-carbon limiting conditions through methanol addition. In this study the pump for methanol addition to reactor C2 was set to deliver a concentration of 70 mg/ℓ as COD.



Note: F refers to feed, RS to return sludge, DENIT to denitrification, PHO RE to phosphorus release, PHO UP to phosphorus uptake, CO to carbonaceous oxidation, NIT to nitrification and ML to mixed liquor

FIGURE 3. PILOT PLANT SINGLE SLUDGE SYSTEMS

Variable speed pumps delivered degritted municipal wastewater at a controlled temperature to each parallel system at constant flows from 5.2 to 8.2 m^3/d (0.8 to 1.25 lgpm) depending on the experiment in progress. The wastewater from the Burlington Skyway Sewage Treatment Plant, exhibited normal diurnal variations in organic and inorganic concentrations (Table 1). The systems consisted of complete-mix stirred and/or dispersed air aeration tanks followed by upflow circular clarifiers for solid-liquid separation. The volumes of the stirred and aerated tanks could be adjusted by varying the liquid heights. Throughout the study the aerobic HRT (nominal) in the systems (Figure 3, reactors A2, B1 and C3) was maintained at six hours. The HRT (nominal) of reactor C2 of system C (Figure 3) was maintained at 2.33 hours. The HRT of reactors A1 and C1 of systems A and C was varied according to the experimental design. The clarifiers were 0.76 m (2.5 ft) in diameter. The clarifier weir heights could be varied allowing maintenance of a nominal HRT from one to two hours. On the basis of the measured sludge blanket heights, and influent and return sludge flow rates, the retention period of the mixed liquor per pass in the clarifiers average approximately 15 minutes for each system.

The experimental design involved parallel operation of the systems at three levels of temperature, 7.5°, 15°, and 25°C, and two levels of HRT, one and two hours (nominal) for the first stage anoxic-anaerobic reactor of sytems A and C (Figure 3, A1 and C1). The aerobic reactor SRT (θ_1) was maintained identical in all three systems during each experimental run in order to compare the effect of anoxic and anaerobic conditions on nitrification. The values of θ_1 maintained were approximately 3, 6 and 14 days for the corresponding temperature levels of 25°, 15°, and 7.5°C. The values are close to the minimum values required to ensure nitrification in com-

TABLE 1. RAW WASTEWATER CHARACTERISTICS

Characteristic (mg/ℓ)	Mean	90%*
COD	286	386
Filterable COD	116	150
BOD_5	106	156
Total P	4.7	6.4
NH_4-N	16.0	18.4
Filterable TKN	20.3	23.3
Alkalinity (as $CaCO_3$)	250	290
SS	164	210

* 90 percent of observations are equal to or less than stated value

pletely aerobic systems[1]. θ_1 was determined from the fraction of suspended solids under aerobic conditions multiplied by the system SRT. The system SRT (θ) was defined as the total volatile suspended solids in the reactors divided by the daily volatile suspended solids lost in the effluent or intentionally wasted. From system B the system SRT maintained at each temperature level was equivalent to the aerobic SRT. For systems A and C the system SRT's maintained were approximately 4, 8 and 18 days and 5, 10, and 23 days, respectively, for the corresponding temperature levels of 25°, 15°, and 7.5°C. The actual value of θ was dependent on the HRT level maintained in reactors A1 and C1 (Figure 3, systems A and C).

In addition to the paired data for systems A, B, and C generated by the experimental design, additional paired data were derived for systems A and B under conditions where;

1. the anoxic-anaerobic reactor of system A (Figure 3, reactor A1) was operated at a nominal HRT of three hours and equal aerobic SRT's and temperatures were maintained in systems A and B, and
2. the aerobic reactors of systems A and B were operated at two levels of SRT, under equal HRT conditions and equal temperatures.

Batch experiments were conducted using sludge derived from system C (Figure 3) to determine the effect of longer anaerobic periods on the degree of P uptake. These experiments also allowed an assessment of the effects of longer anaerobic periods on nitrification. During the batch experiments oxidation reduction potential (ORP) measurements were taken using a flow through cell and procedure similar to that described by Shapiro et al[20].

The batch experiments conducted following the exposure of the nitrifying organism to longer periods of anaerobic conditions, involved operation of

three parallel ten-liter reactors. Seven liters of mixed liquor from reactor C3 (Figure 3, system C) were added to each reactor plus three liters of influent wastewater. The reactors were stirred at a rate just sufficient to maintain complete mixing for periods of one, two and four hours. After each period of mixing, approximately 25 mg/ℓ of NH_4-N (as NH_4Cl) and sufficient alkalinity (as $NaHCO_3$) to prevent pH depression during nitrification, were added to each reactor. Aerobic conditions (dissolved oxygen 2 to 4 mg/ℓ) were then maintained by air diffusion.

The performance of the systems (Figure 3) was assessed from analyses of influent and effluent samples derived from hour by hour sampling over a 24-hour period once the equilibrium SRT and temperature conditions, specified by the experimental design, had been achieved. In all experiments a minimum time delay of 1.5 times the change in system SRT and one week per 5°C change in temperature elapsed prior to an investigative period. In addition to the composite sampling, grab samples of the reactor contents were taken for chemical analyses. These grab samples were filtered immediately. Details of the sample preparation and analytical procedures have been presented elsewhere[19].

The system SRT was controlled automatically by wasting from the return sludge lines. The dissolved oxygen concentration was maintained in the aerobic reactors (Figure 3, A2, B1 and C3) at or above 2.0 mg/ℓ using air diffusion. Reactors A1, C1, and C2 (Figure 3) were stirred at a rate just sufficient to maintain complete mixing conditions.

DISCUSSION OF RESULTS

Effect of Anoxic and Anaerobic Conditions on Phosphorus Removal

The batch and continuous experiments allowed an assessment of how various factors affect the degree of P removal in single sludge systems. The effluent P results from the continuous experiments over the complete range of temperatures and anoxic-anaerobic HRT's in reactors A1 and C1 (Figure 3, systems A and C), indicated an increase in P removal in system C when compared to system B (Figure 4). The results indicated no increase in phosphorus removal in system A when compared to system B (Figure 4). A statistical paired data analysis (Table 2) verified the observed results. The analysis (Snedecor and Cochran, 1967)[21] involved determining the mean difference ($\bar{D}$) between the unfiltered influent P and filtered effluent P observed in Systems A and B, and assessing its statistical significance (Table 2). The resulting "t" statistic was not significant at either a one or two hour anoxic-anaerobic reactor HRT for system A (Figure 3, reactor A1) indicating that no enhanced P removal was observed in system A.

The observed increase in P removal in system C may be due to enhanced biological P uptake or related to an increase in the amount of organic carbon assimilated, due to the addition of methanol to reactor C2, (Figure 3, system C). In order to compare the degree of P removal in systems C and B, the ratio of phosphorus removal to COD removal was calculated for each experimental run. The phosphorus and COD removals were calculated on the basis of the influent unfiltered results minus the effluent filtered results. For system C, the influent COD included that portion contributed by the methanol added to reactor C2 (Figure 3, system C). The results (Figure 5) indicate that the increase in P removal observed in system C was related to

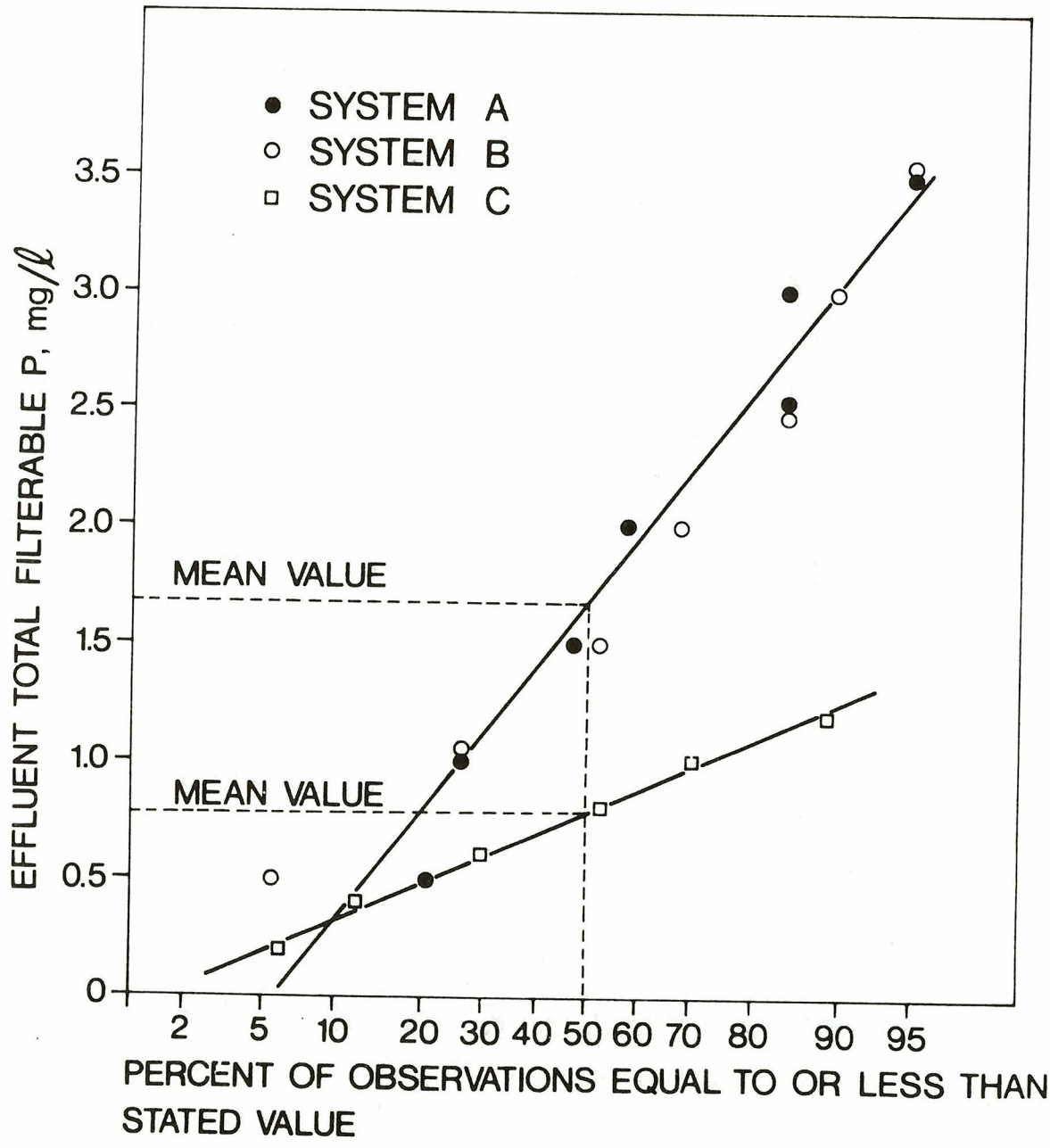


FIGURE 4. EFFLUENT PHOSPHORUS RESULTS FROM AEROBIC AND ANOXIC-ANAEROBIC-AEROBIC SYSTEMS

TABLE 2. DIFFERENCE IN PHOSPHORUS REMOVAL IN SYSTEM A AND SYSTEM B

Al Reactor HRT* (h)	$\bar{D}$	S_D^2/n	t	$t_{\alpha = 0.95}$
1.0	-0.271	0.017	2.11	2.45
2.0	-0.025	0.050	0.11	2.37

* HRT based on influent flow rate only

$\frac{S_{\bar{D}}^2}{n}$ = estimated variance of $\bar{D}$, where

$S_D^2 = \frac{\Sigma(Di-\bar{D})^2}{n-1}$

Di = difference in phosphorus removal (unfiltered influent P - filtered effluent P) between system A and B (P removal A - P removal B)

$\bar{D}$ = mean difference

n = number of paired data points, and

$t = \frac{\bar{D}}{S_D/(n)^{\frac{1}{2}}}$

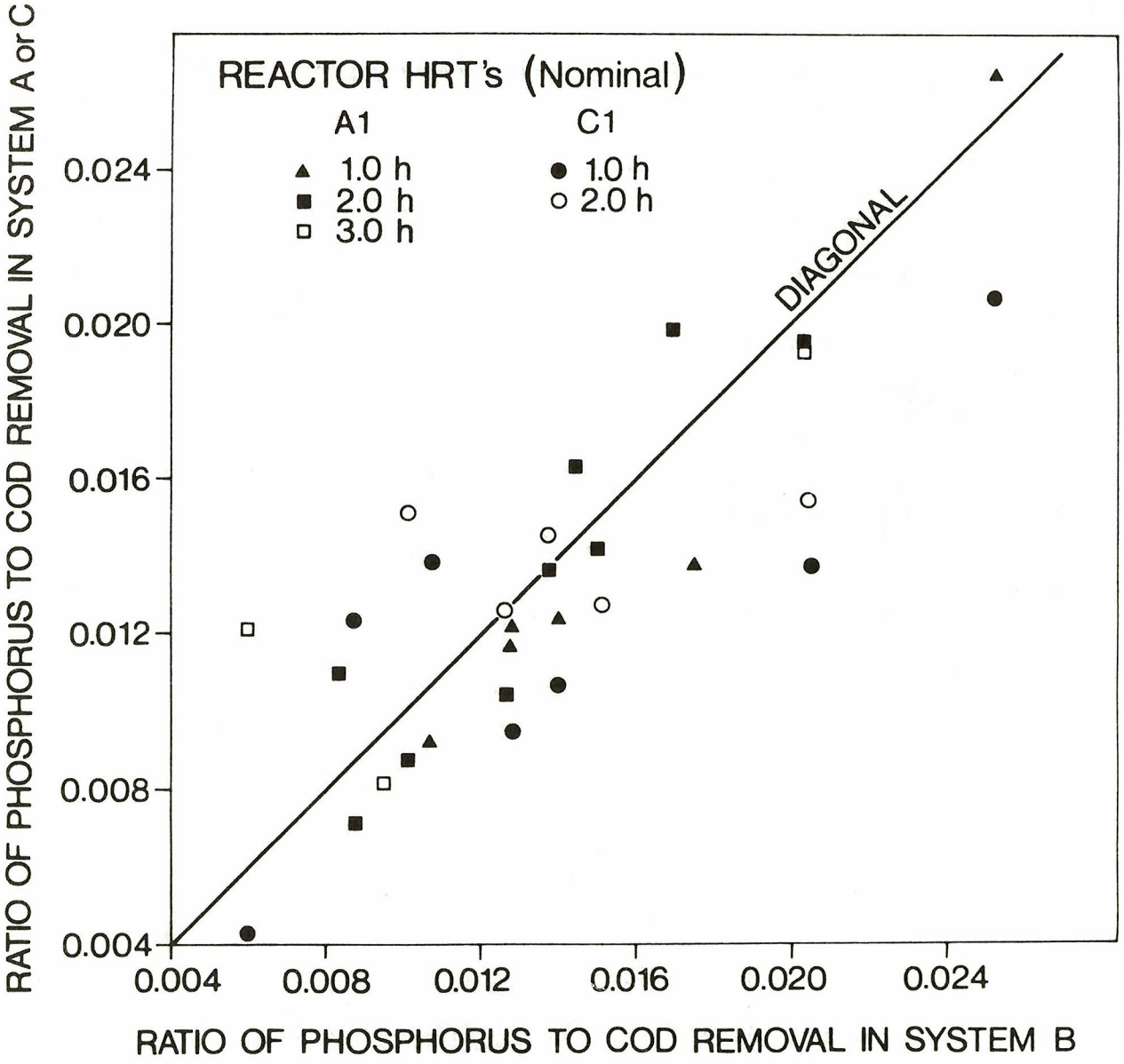


FIGURE 5. PAIRED DATA COMPARISON OF RATIO OF PHOSPHORUS TO COD REMOVAL IN AEROBIC AND ANOXIC-ANAEROBIC-AEROBIC SYSTEMS

an increase in the organic carbon assimilated. The maximum $\Delta P/\Delta COD$ results measured for systems A, B, and C during any experimental run were recorded during parallel operation of the systems at 15°C. The results for systems A and C were 0.027 and 0.021 (Figure 5). The results from system B were 0.025 (Figure 5). The system SRT's were 7, 6, and 9 days for systems A, B, and C, respectively, during this experimental period.

The nitrogen and phosphorus removal performance of configurations D and E (Figure 3) was assessed during a previous study[22]. The results indicated an increase in P removal in the systems when complete nitrate removal was achieved in the denitrification reactors. Calculating the corresponding COD removals it can be shown that the increase in P removal was related to an increase in the concentration of COD fed to the system and a subsequent increase in COD removal. Marsden and Marais[16] report enhanced biological P uptake in the pre-denitrification nitrification sequence (Figure 1, configuration 3) and measured a maximum $\Delta P/\Delta COD$ of 0.022. The temperature and system SRT were respectively 20°C and ten days. Simpkins and McLaren[10] reported enhanced P removal in the "modified" Bardenpho configuration (Figure 1, configuration 2) and measured corresponding $\Delta P/\Delta COD$ ratios of 0.020. The temperature ranged between 17° and 26°C and the system SRT was 15 days in their study.

The lack of an observed biological P uptake in the systems incorporating a first stage anoxic-anaerobic reactor (systems A and C) in this study, was verified from the results. A plot of SRT versus the P content of the mixed liquor in the aerobic reactors of systems A and C (Figure 3, reactors A2 and C3) and of system B (Figure 6) indicates that there is no appreciable change in the P content of the aerobic reactor solids.

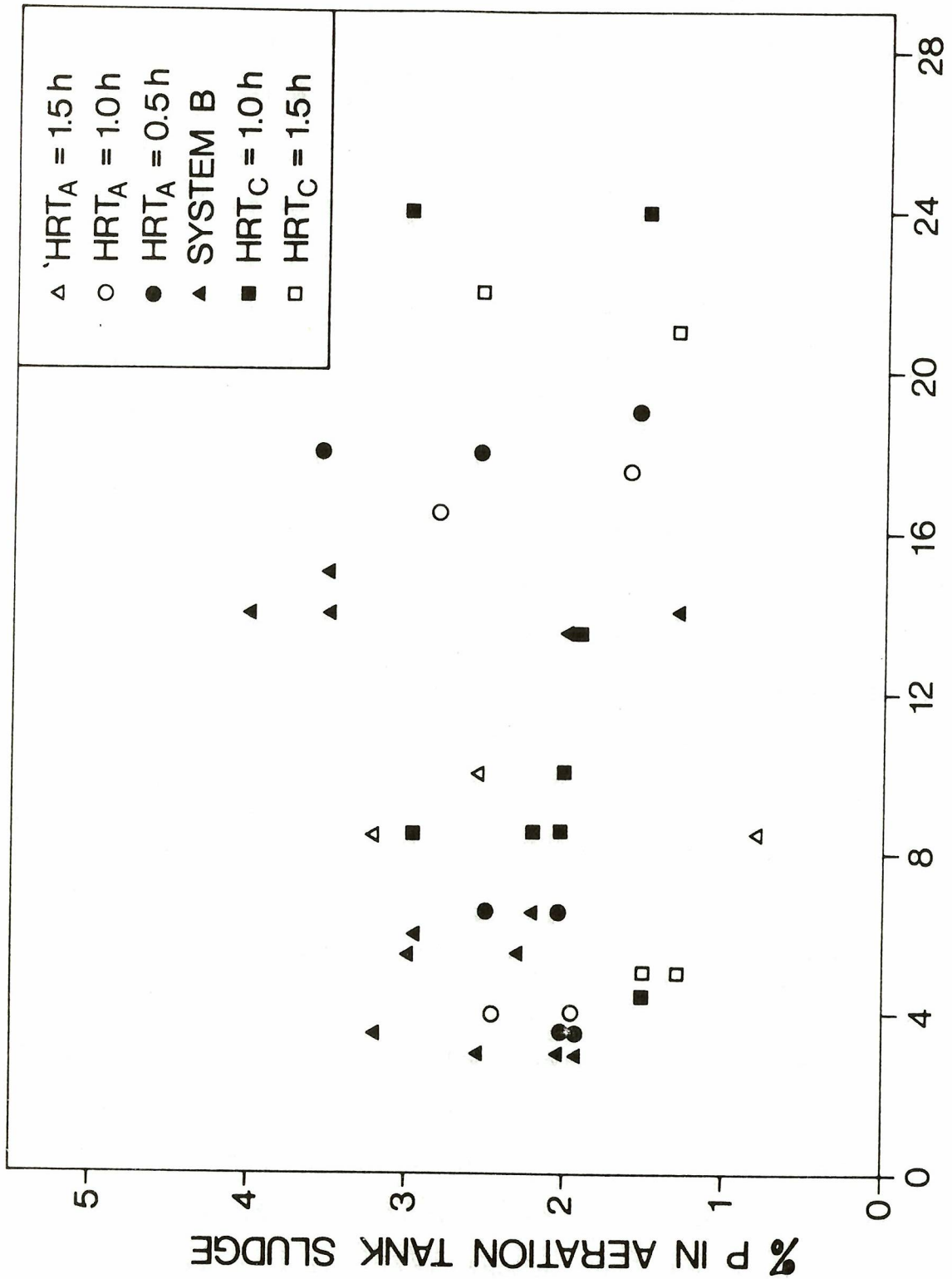


FIGURE 6. PHOSPHORUS CONTENT OF MIXED LIQUOR IN AEROBIC AND ANOXIC-ANAEROBIC-AEROBIC SYSTEMS

The batch experiments allowed an assessment of whether the degree of P uptake is increased by providing longer periods of initial anaerobic conditions. The three parallel batch reactors were operated for periods of 1, 2 and 4 hours under anaerobic conditions followed by an aerobic period of four hours. The filtered total phosphorus results (Table 3) indicate no significant increase in phosphorus removal after aeration when longer periods of initial anaerobic conditions were maintained. The oxidation reduction potential in batch reactor 3 after the four hour anaerobic period was -130 mv verifying that anaerobic conditions had been achieved. The degree of P release during the anaerobic period was significantly less than that observed by Nicholls in his studies of various single sludge configurations[13].

Effect of Anoxic and Anaerobic Conditions on Nitrification

In order to assess the nitrification performance of the systems, both filterable TKN and NH_4-N were determined. The results indicated that filterable TKN removal should be used as the performance measure. In comparing the results for systems A and B (Figure 7), a significantly greater amount of filterable TKN removal was observed in each system. Similar results were observed in system C. It may be expected that a large fraction of the organic nitrogen in the influent wastewater is hydrolyzed to ammonia in the systems and becomes available for nitrification.

The nitrification performance of systems A and B was assessed using the paired data from the experimental design plus the data generated during operation of the anoxic-anaerobic reactor of system A (Figure 3, reactor A1) at a nominal HRT of three hours. The results (Figure 8) indicate that the anoxic-anaerobic conditions had no effect on the growth rates of the nitrifiers regardless of the system temperature (Figure 9). A statistical paired data

TABLE 3. PHOSPHORUS RELEASE AND UPTAKE RESULTS FROM BATCH EXPERIMENTS

Batch Reactor*	Anaerobic Time Period (h)	Aerobic Time Period (h)	Filtered Total Phosphorus (mg/ℓ)		
			Previous to Anaerobic Period	After Anaerobic Period	After Aerobic Period
1	1	4	2.6	4.0	2.9
2	2	4	2.7	5.1	1.7
3	4	4	2.9	4.4	2.8

* Batch tests all at 25°C and using mixed liquor from reactor C3, operating at 25°C and aerobic SRT of three days

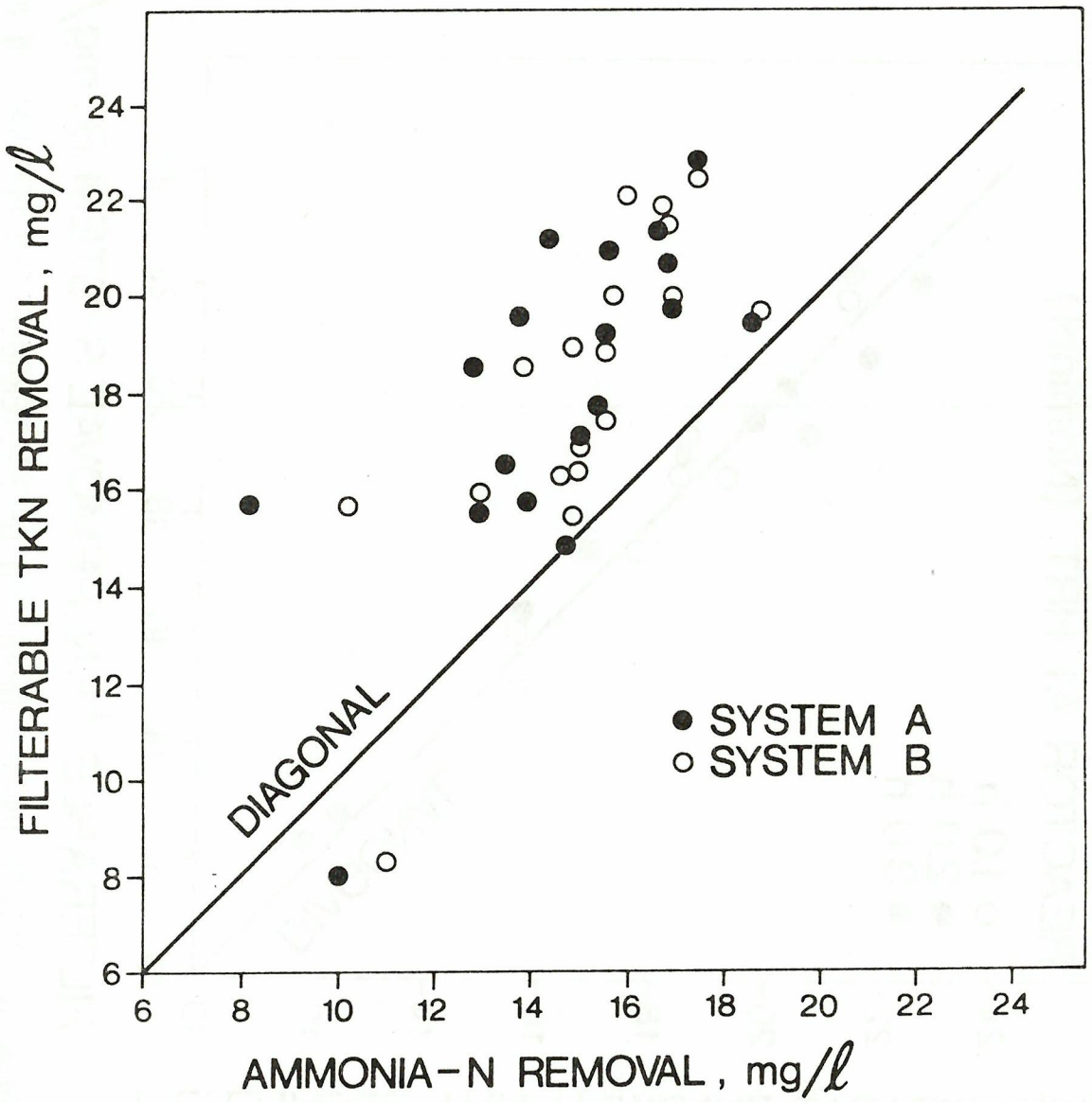


FIGURE 7. AMMONIA-N AND FILTERABLE TKN REMOVAL IN AEROBIC AND ANOXIC-ANAEROBIC-AEROBIC SYSTEMS

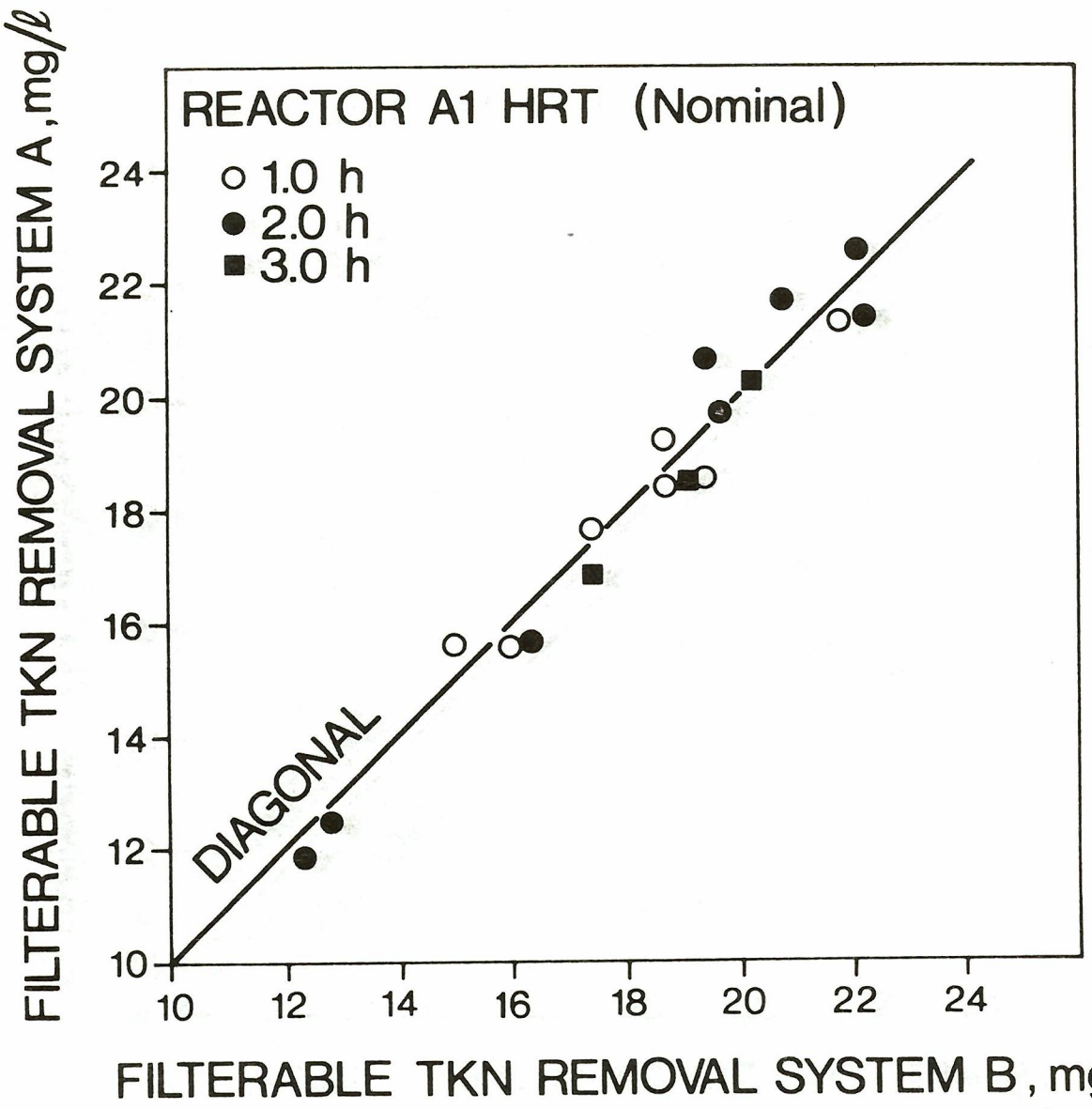


FIGURE 8. PAIRED DATA COMPARISON OF FILTERABLE TKN REMOVAL IN AEROBIC AND ANOXIC-ANAEROBIC-AEROBIC SYSTEMS

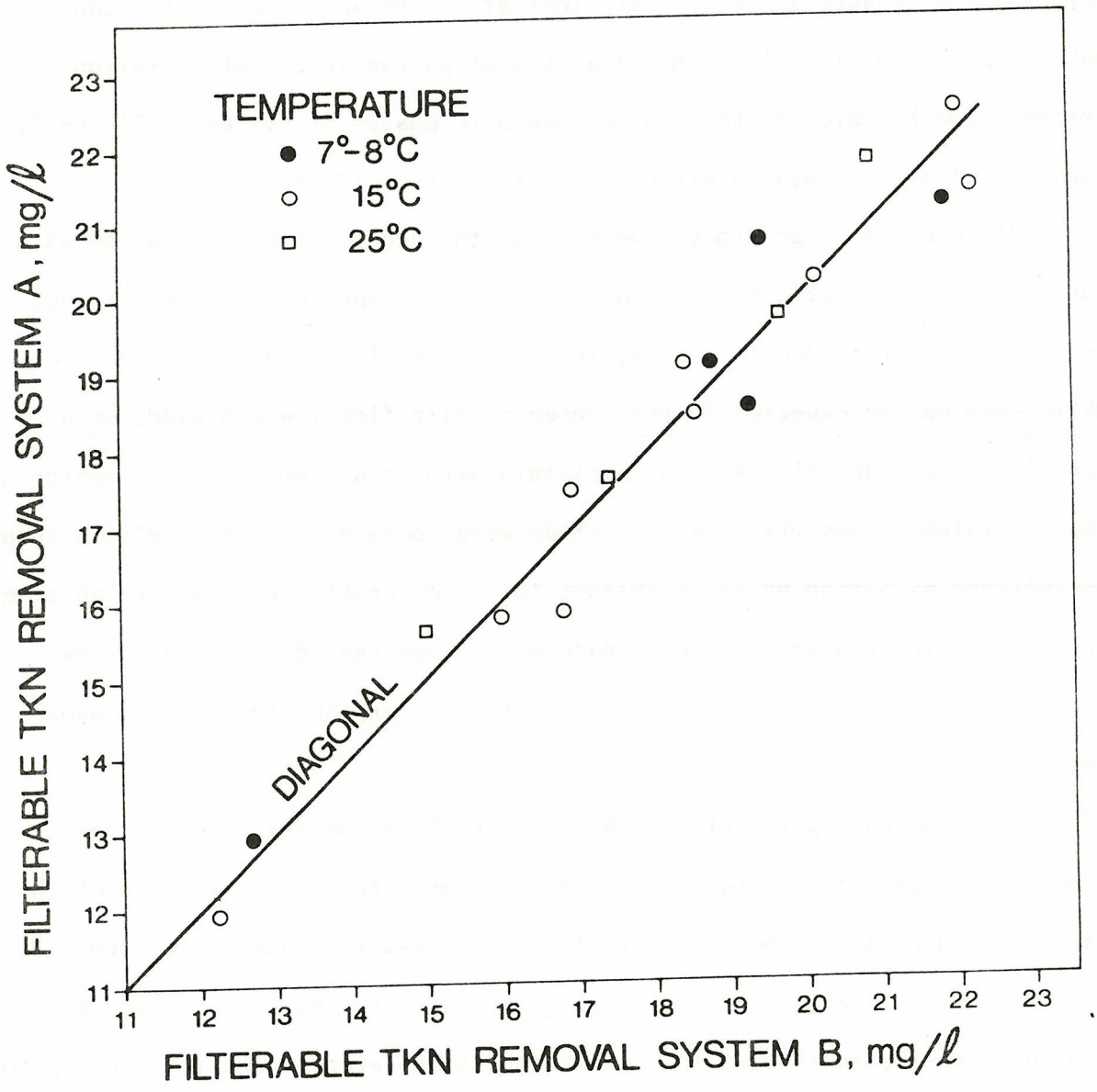


FIGURE 9. PAIRED DATA COMPARISON OF FILTERABLE TKN REMOVAL BY SYSTEMS OPERATING OVER A RANGE OF TEMPERATURES

analysis (Table 4) verified the observed results. In comparing the filterable TKN removal observed in systems A and B, the resulting "t" statistic was not significant at either a one or two hour anoxic-anaerobic reactor HRT for system A (Figure 3, reactor A1) indicating the anoxic-anaerobic conditions had no effect on nitrification. Analysis of paired data from operation of system C and B indicated the anoxic-anaerobic phases of system C (Figure 3, reactors C1 and C2) had no effect on nitrification (Table 4).

Sutton et al.[22] previously determined the degree of nitrification as a function of the aerobic reactor SRT which can be expected in single sludge aerobic-anoxic nitrogen removal systems. The results[22] indicated that no difference may be expected in the degree of nitrification achieved, as a function of the aerobic SRT, in completely aerobic systems and in aerobic-anoxic systems. Results from this study were compared with the relationships established by Sutton et al.[22] (Figure 10). Comparable results were observed (Figure 10) in this study, i.e., there was no apparent difference in the degree of nitrification when the nitrifiers were exposed to aerobic, anoxic, and anaerobic conditions.

In the batch experiments the NO_3-N plus NO_2-N concentration in each reactor was reduced to 0 mg/ℓ within 30 minutes after the initiation of the batch mixing period. The batch nitrification results, once aerobic conditions were established (Figure 11, Table 5), together with results from the continuous experiments (Figure 10), indicate that anaerobic conditions of up to four hours have no effect on the growth rate of the nitrifying organisms.

The single sludge process configurations operated during this study were all designed to achieve complete nitrification. In addition, configurations A and C (Figure 3) were designed to achieve partial nitrogen removal. The

TABLE 4. DIFFERENCE IN FILTERABLE TKN REMOVAL IN PARALLEL OPERATING SYSTEMS WITH AND WITHOUT ANOXIC-ANAEROBIC REACTOR(S)

System Compared to System B		$\bar{D}$	S_D^2/n	t	t $\alpha = .95$
System A					
A1 Reactor HRT*	1.0 h	-0.130	0.039	0.65	2.45
	2.0 h	0.113	0.061	0.45	2.37
System C					
C1 Reactor HRT*	1.0 h	0.129	0.137	0.34	2.45
	2.0 h	-0.780	0.067	2.24	2.78

* HRT based on influent flow rate only

$\frac{S_D^2}{n}$ = estimated variance of $\bar{D}$, where

$$S_D^2 = \frac{\Sigma(Di-\bar{D})^2}{n-1}$$

Di = difference in filterable TKN removal between system A or C and system B (TKN removal A or C - TKN removal B)

$\bar{D}$ = mean difference

n = number of paired data points, and

$$t = \frac{\bar{D}}{S_D/(n)^{\frac{1}{2}}}$$

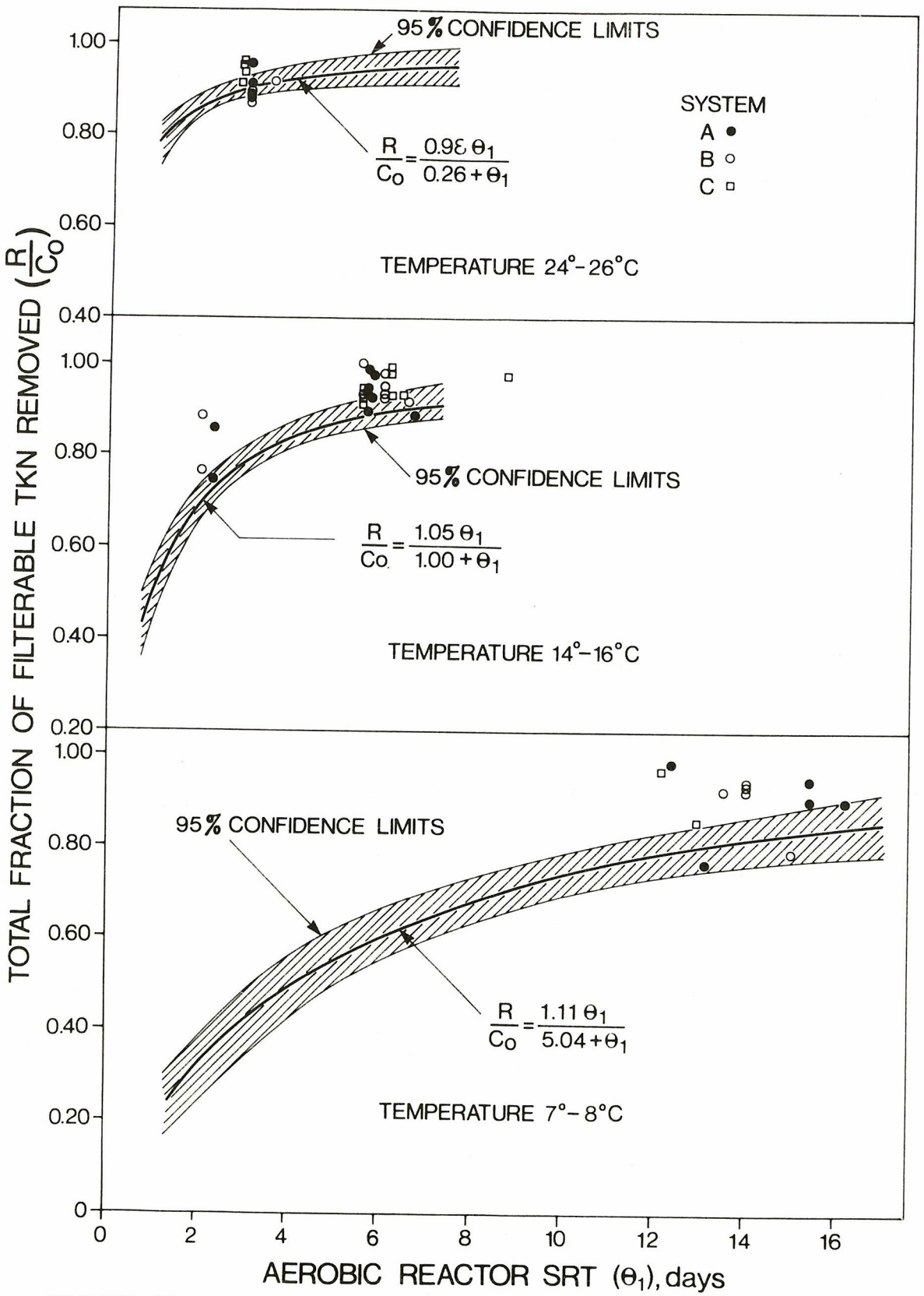


FIGURE 10. COMPARISON OF NITRIFICATION PERFORMANCE WITH PREVIOUSLY REPORTED RELATIONSHIPS[22]

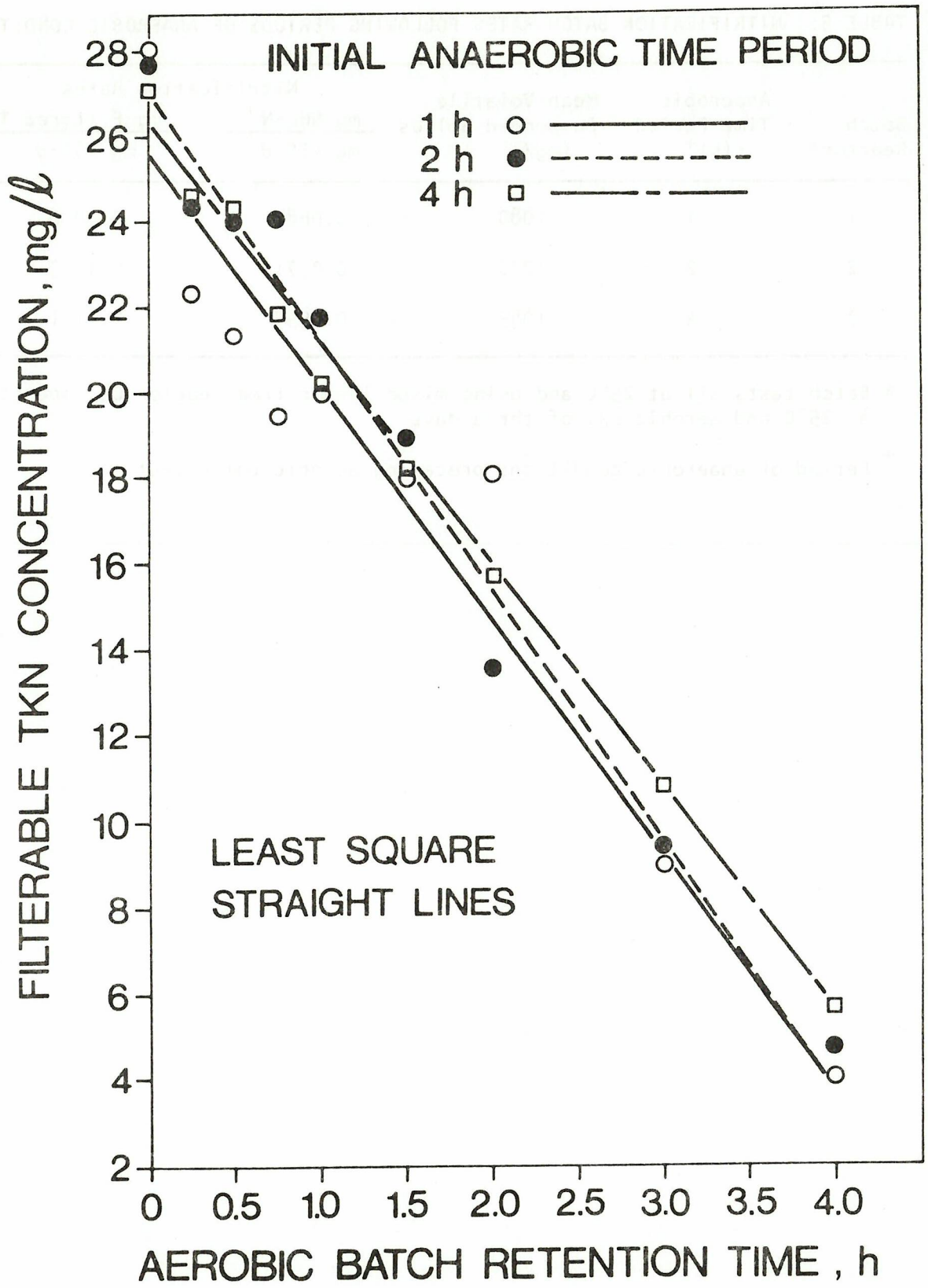


FIGURE 11. BATCH NITRIFICATION RESULTS FOLLOWING PERIODS OF ANAEROBIC CONDITIONS

TABLE 5. NITRIFICATION BATCH RATES FOLLOWING PERIODS OF ANAEROBIC CONDITIONS

Batch Reactor*	Anaerobic Time Period (h)+	Mean Volatile Suspended Solids (mg/ℓ)	Nitrification Rates mg NH_4-N / mg VSS·d	Nitrification Rates mg Filtered TKN / mg VSS·d
1	1	1080	0.0682	0.1076
2	2	1030	0.0771	0.1323
3	4	1045	0.0997	0.1141

* Batch tests all at 25°C and using mixed liquor from reactor C3, operating at 25°C and aerobic SRT of three days

\+ Period of anaerobic conditions preceding aerobic batch test

average filtered N removal obtained in systems A and C was 49 and 82 percent, respectively. From a mass balance around the biological reactors of systems A and C it can be shown that the theoretical filtered nitrogen removal is 50 and 80 percent, respectively. This assumes complete nitrification and denitrification in the corresponding aerobic and anoxic reactors and assumes that all the influent filtered N is available for nitrification.

Methanol was added to reactor C2 of system C (Figure 3) to ensure that the rate of denitrification proceeded under non-carbon limiting conditions. Complete nitrate removal can be achieved in this reactor without methanol addition (Figure 3, system E). In this case the organics present in the raw sewage together with endogenous respiration of the biological sludge, provide the organic carbon necessary for nitrate reduction. The results from system E (Figure 3) derived during previous studies and presented elsewhere[19,20], can be used to determine the conditions necessary to achieve complete nitrate reduction under these conditions.

CONCLUSIONS

The results from operation of pilot plant single sludge systems for nitrogen and phosphorus removal indicate that:

1. Anaerobic conditions of up to four hours had no detrimental effect on nitrification.
2. A high degree of P removal can be achieved in single sludge systems without chemical addition.
3. The amount of P removal is dependent on the concentration of COD removed in the system.
4. Exposing the biological population to alternating anaerobic-aerobic conditions did not promote enhanced P uptake into the sludge.

5. A high degree of N removal can be achieved in single sludge systems without the addition of an external carbon source.

ACKNOWLEDGEMENTS

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REFERENCES

1. U.S.-EPA, "Process Design Manual for Nitrogen Control". Office of Technology Transfer, Washington, D.C. (1975).

2. U.S.-EPA, "Process Design Manual for Phosphorus Removal". Office of Technology Transfer, Washington, D.C. (1976).

3. Barnard, J.L., "A Review of Biological Phosphorus Removal in the Activated Sludge Process", Water SA, 2, 136 (1976).

4. Levin, G.V., et al, "Pilot Plant Tests of a Phosphate Removal Process". Jour. Water Poll. Control Fed., 44, 1940 (1972).

5. Peirano, L.E., "Low Cost Phosphorus Removal at Reno-Sparks, Nevada". Jour. Water Poll. Control Fed., 49, 568 (1977).

6. Price, R.L., and Regan, R.W., "The Fate of Phosphorus in the Phostrip Process". Presented at the 50th Annual Water Pollution Control Federation Conference, Philadelphia, Pa., October (1977).

7. Levin, G.V., et al, "Operation of Full-Scale Biological Phosphorus Removal Plant". Jour. Water Poll. Control Fed., 47, 577 (1975).

8. Campbell, T.L., et al, "Biological Phosphorus Removal at Brockton, Massachusetts". Jour. New England Water Poll. Control Assoc., 12, 16, (1978).

9. McLaren, A.R., and Wood, R.J., "Effective Phosphorus Removal From Sewage by Biological Means". Water SA, 2, 47 (1976).

10. Simpkins, M.J., and McLaren, A.R., "Consistent Biological Phosphate and Nitrate Removal in an Activated Sludge Plant". Progress in Water Technology, 10, (5/6), 433 (1978).

11. Osborn, D.W. and Nicholls, H.A., "Optimization of the Activated Sludge Process for the Biological Removal of Phosphorus". Progress in Water Technology, 10, (1/2), 261 (1978).

12. Venter, S.L.V., et al., "Optimization of the Johannesburg Olifantsvlei Extended Aeration Plant for Phosphorus Removal". Progress in Water Technology, 10, (1/2), 279 (1978).

13. Nicholls, H.A., "Kinetics of Phosphorus Transformations in Aerobic and Anaerobic Environments". Presented at Post Conference Seminar PSI of the 9th IAWPR Conference, Stockholm, Sweden, June (1978).

14. Martin, K.A.C., and Marais, G.v.R., "Kinetics of Enhanced Phosphorus Removal in the Activated Sludge Process". Research Report No. W14, Dept. of Civil Eng., University of Cape Town, (1975).

15. Stern, L.J. and Marais, G.v.R., "Sewage as Electron Donor in Biological Denitrification". Research Report No. W7, Dept. of Civil Eng., University of Cape Town, (1974).

16. Marsden, M., and Marais, G.v.R., "Role of the Primary Anoxic Reactor in Denitrification and Biological Phosphorus Removal". Research Report No. W19, Dept. of Civil Eng., University of Cape Town, (1976).

17. Hoffman, R.J., and Marais, G.v.R., "Phosphorus Removal in the Modified Activated Sludge Process". Research Report No. W22, Dept. of Civil Eng., University of Cape Town, (1977).

18. Wuhrmann, K., "Effect of Oxygen Tension on Biological Treatment Processes". Proceedings 3rd Conference on Biological Waste Treatment, Manhattan College, New York, U.S.A., p. 27 (1960).

19. Sutton, P.M., et al, "Single Sludge Nitrogen Removal Systems", Canada-Ontario Agreement Research Report No. 88 (1979).

20. Shapiro, J., et al., "Anoxically Induced Release of Phosphate in Wastewater Treatment". Jour. Water Poll. Control Fed., 39, 1810 (1967).

21. Snedecor, G.W., and Cochran, W.G., "Statistical Methods", Iowa State Univ. Press, Ames, Iowa, p. 115 (1967).

22. Sutton, P.M., et al, "Design Considerations for Integrated Nutrient Removal Systems", Progress in Water Technology, 10, (5/6), 469 (1978).

LAND APPLICATION OF SECONDARY EFFLUENT

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PREPARED FOR

NUTRIENT REMOVAL SEMINAR
CALGARY, ALBERTA

FEBRUARY 7 AND 8, 1980.

Vernon
Cranbrook } irrig. with eff.

Muskegon, Michigan
Kelowna - Bardenpho → then eff. irrigation
- trickle method.

LAND APPLICATION OF SECONDARY EFFLUENTS

Land applications of secondary effluent can be achieved by the use of many techniques. The use of soil as both a treatment and disposal medium is accomplished by means of perforated underground pipe or by the use of ex-filtration ponds. In such cases, there is little or no use made of the water or any impurities therein, and hence sufficient unsaturated soil must be present to act as a treatment medium for the economic life of the installation.

Other common forms of land application generally provide a certain degree of resource recovery, the most obvious example being the utilization of both water and nutrients for crop production. The methods of application that are possible include spray irrigation, ridge and furrow irrigation, flood irrigation, and trickle or drip irrigation. The choice of systems depends upon such variables as topography, type of crop, and cost of water.

When some form of land disposal is being considered for a given community, it is essential that available sites be adequately investigated before any irrevocable decisions are made. It is quite possible that some municip-alities do not have sites within an economically viable distance that are satisfactory for the purpose. In such cases, extra treatment plus water disposal may be indicated. It is imperative to recognize that the land cannot be arbitrarily used for waste disposal without the potential being present for serious problems to occur.

POTENTIAL HEALTH EFFECTS

There are a variety of impurities that may be present in wastewater and that may have detrimental effects on human health when such wastewaters are discharged on land. In this regard, however, one should keep in mind the relative danger to human health from land disposal of wastewater vs. that from water disposal of the same wastewater.

In considering domestic wastes, the impurities of measurable concern to human health are biological in nature, and can be grouped into the following categories:

1. bacteria
2. viruses
3. intestinal parasites
4. amoeba

Some of the more common diseases that have, on occasion, been traced to fecal origin are as follows.

bacteria -	bacillary dysentery
	cholera
	typhoid fever
	paratyphoid fever
	tuberculosis
	salmonella gastroenteritis
virus -	infectious hepatitis
	poliomyelitis
	viral gastroenteritis

parasites -	hookworm
	whipworm
	beef tapeworm
	pork tapeworm
	ascariasis
	enterobiasis
amoeba -	amoebic dysentery

The transmission of the disease organisms through the medium of sewage can be grouped into three categories when considering land disposal.

1. direct human contact with sewage
2. ingestion of contaminated food crops
3. presence of an intermediate host

DIRECT HUMAN CONTACT WITH SEWAGE - The danger of disease transmission by this route is real, and must be guarded against. Workers should practise essentially the same sanitary habits as recommended for sewage treatment plant operators. However, it should be kept in mind that the degree of danger to irrigation workers will be substantially less than to sewage treatment plant operators, because the former are dealing with treated sewage only. Furthermore, quantification of health dangers to even sewage treatment plant operators has been very difficult to establish, and hence the value of specific safety programs cannot be established either.

Regardless of the difficulties involved in establishing a causal relationship between direct contact with sewage and incidence of disease in Canada or the USA, there is no doubt that adequate precautions should be taken to minimize public contact with fresh or even secondary treated sewage. The

possibility of contracting a disease from such contact is of course governed by both statistics and susceptibility - the greater the number of disease organisms present in a given quantity of water, the greater the chance of infection. In Canada and the USA, the general level of infection in the population is extremely low, and hence the concentration of disease organisms in sewage is also relatively low. It follows that if we slacken off on our precautions against disease spread by way of contact with sewage, the incidence of such disease in the human population can slowly grow - and as it slowly grows, the concentration of disease organisms in sewage will grow, and hence the statistical chance of infection from contact with sewage will in turn grow. It therefore becomes a vicious circle.

This very thing has been allowed to happen in Mexico. Data from Mexico indicate a nationwide mortality rate from water-borne diseases to be 1.3 per 1000 population in the year 1971. Morbidity rates from the same cause would be at least one order of magnitude higher, and probably more. Even if good water pollution abatement measures are implemented immediately, it will take a considerable length of time to break the cycle of infection and reduce the incidence to Canadian or US levels.

INGESTION OF CONTAMINATED FOOD CROPS There is no doubt that disease transmission has occurred due to the ingestion of food crops that have been contaminated with pathogens from irrigation with sewage. Parsons et al[8] quoted a comprehensive review of such disease outbreaks undertaken by Bryan[3]. That review lists some 63 documented cases throughout the world (primarily North America and Europe) that have occurred since 1894. In Bryan's words "The foods incriminated in these outbreaks were found to be grossly contaminated with night soil or raw sewage".

If the dangers of food contamination arising from irrigation with secondary effluent are considered, the evidence appears to be much more sparse. However, there is sufficient data to show that significant pathogenic and parasitic organisms can and do survive the secondary treatment process, even when normal disinfection practice is followed. It follows, therefore, that there is still some chance that these surviving organisms will contaminate food crops irrigated with the effluent to the extent that disease symptoms can occur in humans.

The method of irrigation, combined with the type of crop, has a great influence on the possibility of food contamination. Ridge and furrow irrigation, for instance, should have little effect on that part of the plant which grows above the ground. Recent bacteriological analyses performed on various food crops in Mexico have shown that tomatoes irrigated with raw sewage (ridge and furrow method) had extremely low coliform counts as gathered by technicians in the field. However, the handling of the fruit by field workers gave rise to appreciable contamination - probably because the field workers' hands became contaminated through contact with the soil. Root crops, of course, would have direct contact with the wastewater regardless of the method of application.

Essentially all of the diseases listed previously could be transmitted via food crops contaminated with sewage containing the causative agent, as all of these agents have finite lifespans outside the human body. A summary prepared by Parsons et al[8] indicated that the fastest die-off rate on vegetables is attributed to cysts of the causative agent for amoebic dysentery, with a time of viability ranging from less than one day to three days. The longest reported survival belongs to the eggs of the roundworm Ascaris, which appear to survive on vegetables and fruits for 30 to 35 days. Incidentally, the survival time in soil is reputed to be up to seven years.

INFECTION VIA INTERMEDIATE HOSTS: There are two fairly well-known examples of parasitic infections of man that involve sewage and intermediate hosts. Beef tapeworm (Taenia saginata), can infect man when he eats poorly cooked beef that contains T. saginata larvae. The larvae change to adult tapeworms in the human host, and eggs or ripe worm segments are excreted in the feces. If the sewage is then applied to crops that cattle eat, the cycle is completed. The pork tapeworm (Taenia solium) can complete its cycle in a similar manner, although it can also be cycled directly from man - feces - man.

AGRICULTURAL IMPLICATIONS OF LAND DISPOSAL OF SEWAGE

The agricultural implications associated with wastewater disposal on land can be divided into two broad categories: soil effects; and crop effects. If it is desired to use a given plot of land for the growing of crops for a long period of time (say, greater than 10 years), then it is very important to understand the effects that impurities in the sewage might immediately have upon the crops, and the long-term effects that might occur due to root-zone soil changes.

EFFECTS ON SOIL: There are a number of characteristics that affect the ability of soil to accept wastewater with no significant deleterious results within the root-zone of crops. For convenience we can group these characteristics into physical, chemical, and biological attributes.

Physical Characteristics

- soil permeability
- suspended solids content of wastewater
- depth of soil unsaturated with water

Chemical Characteristics

- adsorptive capacity of soil
- exchange capacity of soil
- complexing and chelating ability of soil
- dissolved impurity content of wastewater
- depth of aerobic zone
- strength of impurity bonding to soil particles

Biological Characteristics

- impurity uptake by crops
- water uptake by crops
- BOD content of wastewater
- aerobic vs. anaerobic regimes

The ideal situation with regard to soil would be the presence of a great depth of unsaturated soil with a high capacity for permanently extracting pollutants from wastewater. This would mean that heavy metals, refractory organics, and the like would be immobilized on the soil particles, and would not be able to enter the food chain through crop uptake, nor enter the groundwater regime by leaching during the rainy season, or when excess irrigation water is applied. Unfortunately, this ideal situation is seldom encountered, and it therefore becomes important to have a good understanding of the soil regime at any proposed wastewater irrigation site before any facilities are designed.

Phosphorus

In the Okanagan Valley, the soil characteristics vary considerably from fairly clean sand to loamy silt or loamy clay. Some work done by Adriano[1] (as quoted by Webber and Leyshon[10]) showed that somewhat similar soils had phosphorus adsorption capacities as follows:

dune sand - 77 lb. of P per acre-foot

clay loam - 858 to 1898 lb. of P per acre-foot

In work done at Vernon, Oldham and Kennedy[4] found the capacity of a natural Okanagan sand to be 27 lb. of P per acre-foot, with a much higher, although undetermined, value for silty loam. Using this data, and assuming that the normal application rate of 30 inches of treated domestic sewage were applied to the two types of soil each year, the movement of a discrete phosphorus "front" would occur as indicated in Figure I. It is apparent that sewage disposal by irrigation in the sandy areas of the Okanagan, even after secondary treatment, could easily cause the exchange capacity of the soil for phosphorus to be exhausted, if utilization of this nutrient form by crops is not achieved.

Nitrogen:

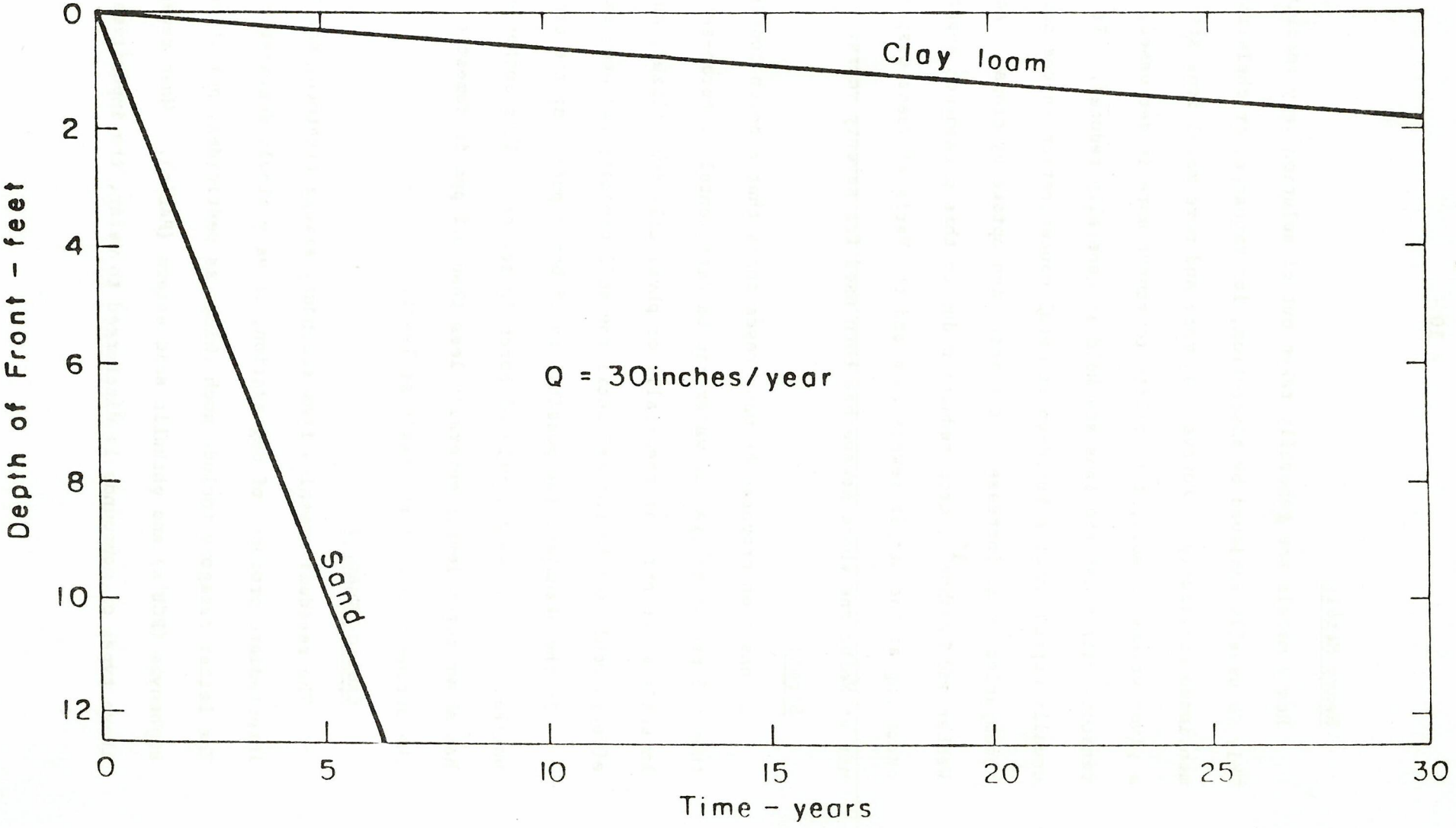
Where municipal sewage (with or without secondary treatment) is used for crop irrigation, there is a significant amount of nitrogen added to the soil. If the irrigation is undertaken at a rate close to the water requirement of a crop such as wheat, about 90% removal of nitrogen[7] is achieved. However, if a leguminous crop such as alfalfa is being raised, the situation is complicated by the production of nitrogen that occurs on the root nodules.

If the soil is aerobic for the full depth of the unsaturated zone, nitrification of the ammonia in sewage occurs, with the highly mobile nitrate ion being formed. If any water in excess of crop requirements is added, the formed nitrate is quickly carried through the root zone by the excess water.

Fig.

ANTICIPATED MOVEMENT OF PHOSPHORUS FRONT THROUGH OKANAGAN SOILS IRRIGATED WITH SEWAGE.

(No crop uptake)



Heavy Metals:

Heavy metals are generally taken out of solution very readily by soils. This is usually achieved by adsorption, ion exchange, or chelation, as mentioned previously. However, as more and more metal ions are removed by a given volume of soil, the ability to remove more is decreased, and the tenacity with which the ions are held is generally reduced. It therefore usually happens that a build-up in metal concentration in the soil is accompanied by an increase in the metal ion uptake by crops. According to Webber and Leyshon[10], crop reduction due to this phenomenon appears to be occurring at the Berlin sewage farm and the Paris effluent disposal site, but not at Melbourne where sewage has been used for seventy years.

Boron:

It has been proposed by many researchers that a boron content greater than 0.5 ppm in irrigation water can be detrimental to boron-sensitive crops. In soils where boron is chemically or physically immobilized, such crop effects would not be noticed until the soil capacity had been exhausted.

In the Okanagan, the possibility of boron poisoning the crops seems unlikely with present levels of boron in sewage. This author's experience has shown boron levels generally less than 0.2 ppm in domestic sewage from the northern part of the basin at least.

Exotic Organics:

The residual organics from secondary sewage treatment are either intermediate products of degradation, or very slowly degradable exotics. The latter category include such things as pesticides, polychlorinated biphenyls (PCB's) and phthalic acid esters (PAE's). When sewage containing these types of compounds is discharged to water, the impurities have a fairly

direct route for entering the food chain and concentrating in higher animals. When discharge is to land, there appears to be a much greater chance of slow degradation of these organics occurring, because soil is an extremely good reservoir for many kinds of bacteria, some of which can almost assuredly decompose most residual organics. In addition, chemical and physical reactions tend to hold the residual organics within the soil structure, thus giving bacterial populations more time to work on the organics.

The main danger would appear to be that of applying an excess of such organic material, thus causing toxicity reactions within the biological community of the soil.

EFFECTS ON CROPS: There are two impurity effects on crops that are of concern:

1) the presence of material in sufficient concentrations to affect the viability of the crop itself;

2) the uptake of impurities that may cause disease or toxic reactions in the population that uses the crops, even though there may not be any detrimental effects on the crops themselves.

The health of the crop itself can be affected by many of those compounds discussed under the section on soils. Uptake of excessive metal ions, boron, and even phosphorus can have a debilitating effect on many crops. In addition, the water itself can detrimentally affect the crops. In Vernon, for instance, it was found that alfalfa was adversely affected by pockets of water forming within the root-zone of the crop. Also where irrigation was carried on into the late Fall, root damage occurred if freezing conditions arrived before a substantial snowfall. If the crops are to be successful

in their role of a resource management tool, their welfare must be provided for in both the design and operation of the irrigation project.

Response of the crops to more or less water, and more or less macro-nutrients (N & P) is important to the design process. Bole[2] has shown that doubling the effluent application rate from 28.5 inches per year to 57.0 inches per year at Taber, Alberta, caused a substantial increase in the yield of Reed Canary Grass, Altai , Wild Rye, Bromegrass, Tall Wheatgrass, and Alfalfa. This information, if taken in isolation, could be interpreted to mean that effluent application at substantially more than the anticipated rate of evapotranspiration yields good results. Three things, however, should be kept in mind:

1) the build-up of impurities in the soil strata will be much faster at the higher application rate;
2) the leaching effect of the high volume of excess water will tend to wash impurities through the unsaturated soil to the groundwater;
3) the crop return from irrigating twice as much land at the lower rate will be much higher. (see Fig.II).

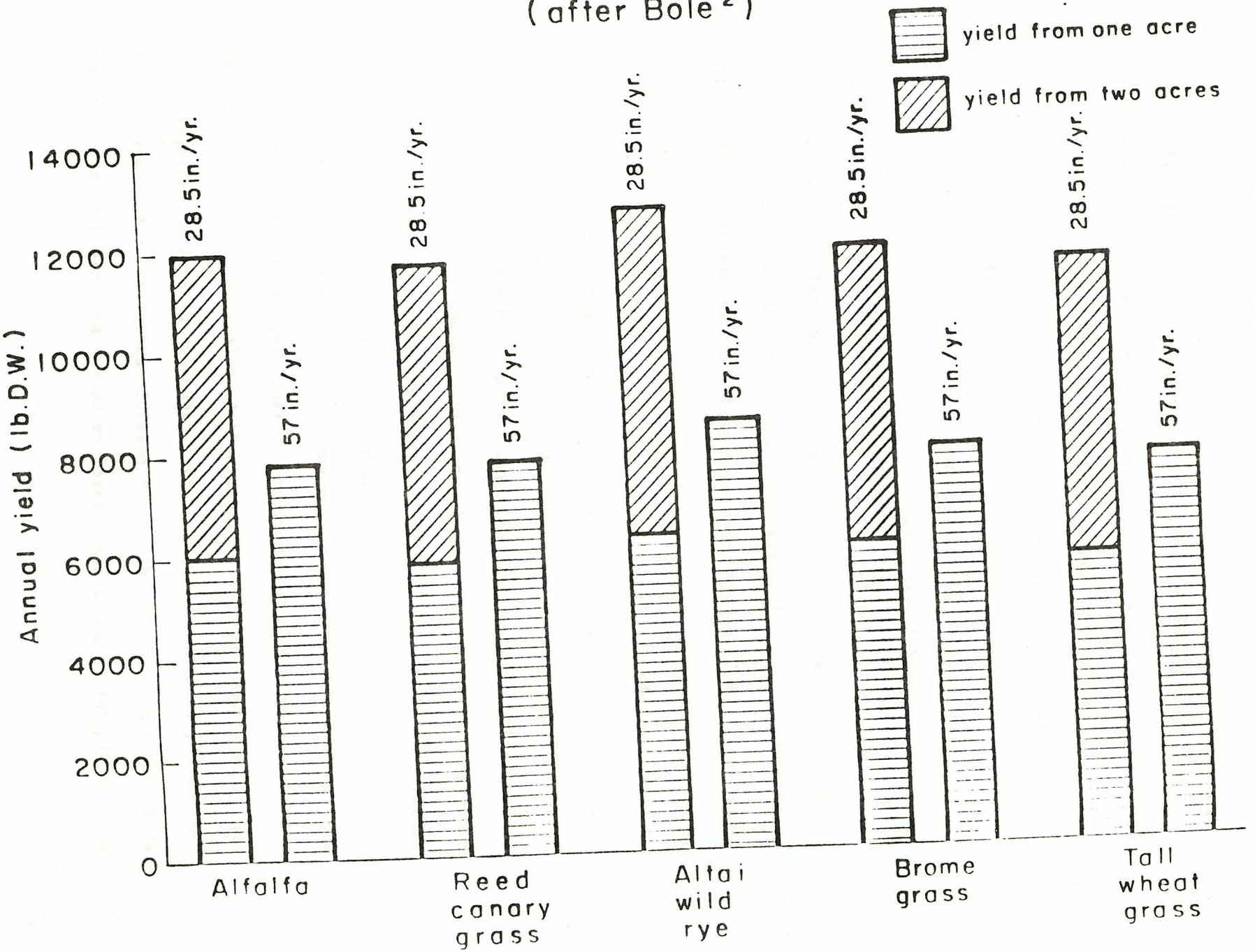
Of course, when short term economics are concerned, item 3) above will be insignificant if the value of that extra yield has to be compared to the annual cost of purchasing the extra land.

The important thing to recognize regarding the growing of crops with effluent, is that a good knowledge of effluent characteristics be achieved, for both beneficial and detrimental impurities.

FI

COMPARATIVE CROP YIELDS FOR DIFFERENT EFFLUENT APPLICATION RATES.

(after Bole[2])



WATER POLLUTION POTENTIAL OF EFFLUENT IRRIGATION

Two conditions must generally be satisfied in order for effluent irrigation to give rise to pollution of groundwater or surface water.

1) The effluent must reach a body of water that is being used, or may be used, for some purpose.

2) The quality of the effluent that finally reaches the body of water being considered must be such that it detrimentally affects the usefulness thereof.

If effluent is applied to the land in excess of the crop requirements for water, that excess liquid must enter the general water resource inventory of the area. The effect that this lost water will have on the receiving water quality will depend upon the impurities that are carried along in solution or suspension.

In general, the impurities of concern to receiving water quality can be grouped into three categories: those that affect public health; those that cause eutrophication; and those that detrimentally affect aquatic life. Public health concerns generally are limited to disease causing organisms, heavy metals, potentially carcinogenic or teratogenic organics, and high nitrate concentrations. Eutrophication problems are almost universally traceable to increased nitrogen and/or phosphorus inputs, while aquatic life is generally affected by the same constituents that affect public health.

The effects of such impurities on receiving waters will generally be much less pronounced when the intermediate effects of land application are considered, than when there is direct discharge of sewage treatment plant effluent to a receiving water. A possible exception to this would occur where exfiltration ponds are located on coarse soil close to a body of water.

POSSIBLE PATHWAYS FOR POLLUTANT TRAVEL TO RECEIVING WATERS: When effluent is applied to land by any means, the water which is not consumed by the crops will be lost from the site, and will carry some impurities with it. In addition, impurities which are extracted by the soil during irrigation, may be partially released during the wet season, or during the snow-melt period, due to a process known as leaching.

Overland Loss During Irrigation: When the rate of effluent application is greater than the rate at which the soil can physically accept the water, overland loss can occur. In such cases, the impurity content is generally lower after the effluent has travelled over the ground because of chemical and biological reactions with the surface soil particles. Thomas[9] and Ehlert[10], for instance, obtained very high removal rates of nutreints and residual BOD with a designed overland treatment system.

This situation may occur inadvertently in a spray irrigation system that has not been adequately designed or operated. Toward the end of an irrigation season it is possible that the surface layer of soil becomes more impermeable due to such things as a poor calcium-sodium balance in the effluent, or a high organic content in the effluent which in turn can lead to soil clogging due to the abundant growth of bacteria that occur.

Overland Runoff During Rainstorms or Snow-melt: The travel of water over the ground surface can cause the release of some impurities that were adsorbed during the irrigation process.

Vertical Leaching By Excess Irrigation Water: Normal irrigation practice demands that the total amount of irrigation liquid applied be greater than the amount actually consumed by the crops. This practice allows highly soluble salts that are deposited in quantities greater than equilibrium amounts to be flushed away from the root zone of the crops, and hence helps to prevent the formation of poor chemical conditions that might hinder crop growth.

The question of the desired amount of excess water is not answerable in terms of a definite percentage that is universally acceptable. In fact, the actual amounts of water used by various crops under various climatic conditions are not well-documented. In general, the more excess water applied, the more impurities that will be moved deeper into the soil horizons, and eventually into the groundwater reservoirs.

Vertical Leaching By Rainfall or Snow-melt: Impurities previously adsorbed onto soil particles can be partially removed by the percolation of relatively pure water from rainfall or snow-melt. This is especially true on irrigated land because the soil is often close to its maximum water retention content at the time the precipitation or snow-melt occurs. Therefore, a significant portion of the new water can pass through the crop root zone, rather than being retained therein for crop use.

POTENTIAL SEVERITY OF PROBLEM: The danger of water pollution resulting from irrigation of secondary effluent is probably the least worrisome aspect of such a disposal practice. As long as adequate consideration is given to soil type and available depth of unsaturated soil at the time of facility design,

and as long as inordinately high liquid loadings are not imposed upon the soil once operation of the facility has begun, water pollution problems will be many times less than those which occur from direct water discharge of the same treated effluent.

DESIGN CONSIDERATIONS

The use of any land disposal system must be preceded by an adequate design phase, if we wish to avoid potentially serious future effects of such use. It is possible, of course, to minimize the potential for future problems by simply using grossly safe design procedures without regard to the variable ability of disposal sites to assimilate impurities. The drawback to such a course of action lies in the fact that the cost will generally be considerably higher than necessary to obtain acceptable environmental conditions in the soil, crops, and receiving waters.

EXFILTRATION BASINS: The prime requisite of exfiltration basins is that they be located in an area which will not give rise to pollution of groundwater or surface water bodies. Sufficient information should be obtained regarding the soil conditions to answer the following:

1) how good is the exchange or adsorptive capacity of the underlying soil?
2) what depth of unsaturated material is present?
3) where do any highly porous strata lead?
4) what distances are involved to the nearest surface water? to the nearest well?
5) what is the average permeability of the underlying material?

SPRAY, RIDGE & FURROW, OR TRICKLE IRRIGATION: For any one of these three methods of irrigation, the basic requirement is that an understanding be achieved regarding the amount of irrigation liquid that is required to just meet the crop's consumptive needs during the growing season. Once this information has been estimated, then various alternative designs can be considered. Two "extremes" in design will be considered herein: 1) optimization of resources, and; 2) maximization of effluent application.

OPTIMIZATION OF RESOURCES: The term "optimization of resources" in this context is taken to mean the use of the water and macro-nutrients present in the effluent to gain as close a balance as possible between amounts irrigated, and amounts removed by the crops. In order to achieve this sort of balance, it is necessary to have a good idea of the tonnage of harvestable crop, the chemical characteristics of that crop when grown on the chosen land with effluent as an irrigation supply, and the net consumptive requirement of the crop for water.

The investigations undertaken at Vernon[5] during 1971 and 1972 were designed to determine such an optimum design for the growing of alfalfa in the Vernon area. Figure III shows the results of a water balance calculation, using both the Department of Agriculture estimate of gross evapotranspiration, and the evapotranspiration calculated from concurrent lysimeter studies at Penticton[6]. The percentage utilization of effluent ranges from about 55% (DoA method) to about 85% (lysimeter method)

Phosphorus balance information for the same year is shown in Figure IV, and indicates that about 72% of the applied phosphorus was removed with the crop. It should be noted here that irrigation in that particular year

FIG. III
1972 WATER QUANTITY BALANCE.
(MAY - OCTOBER)

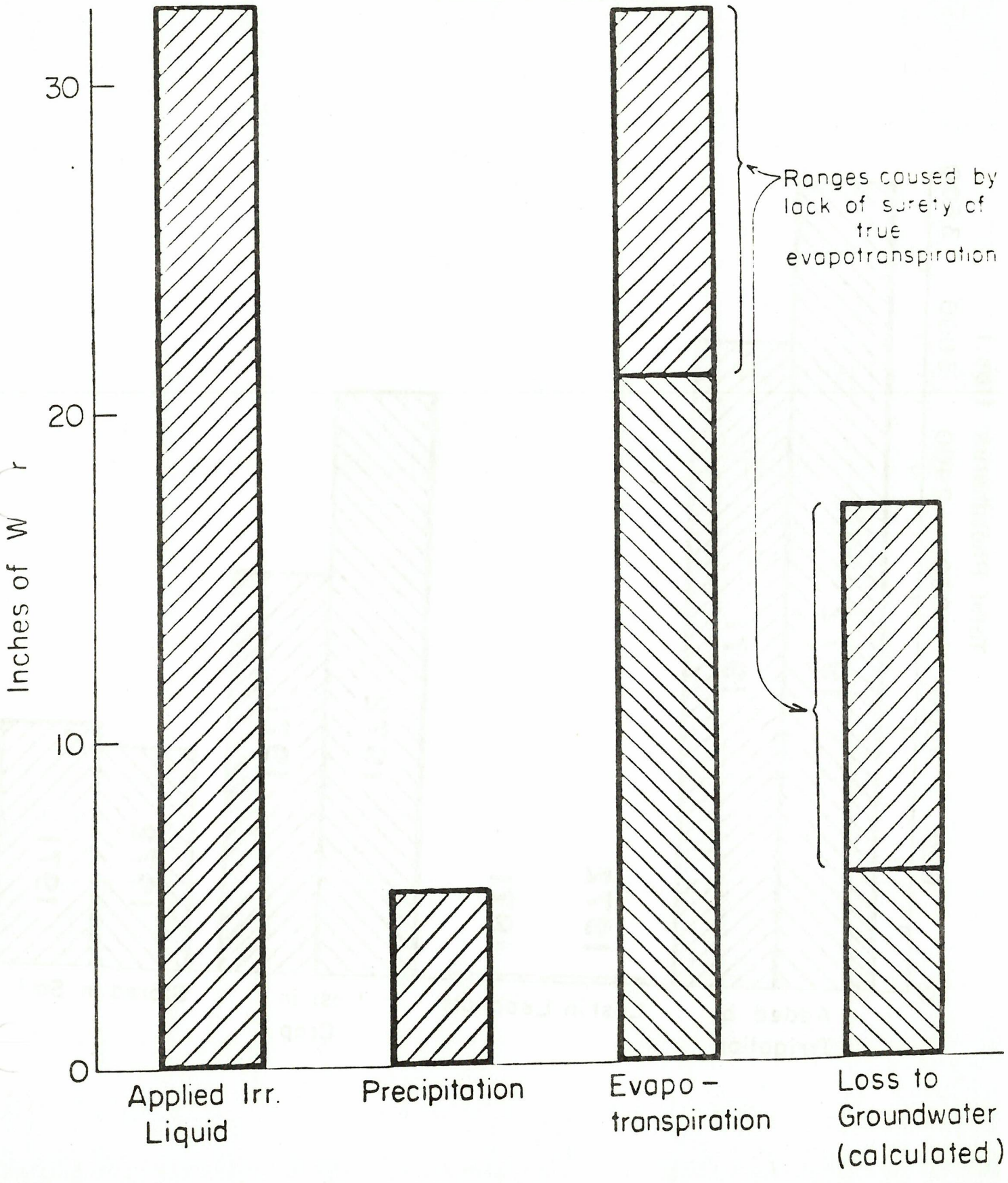
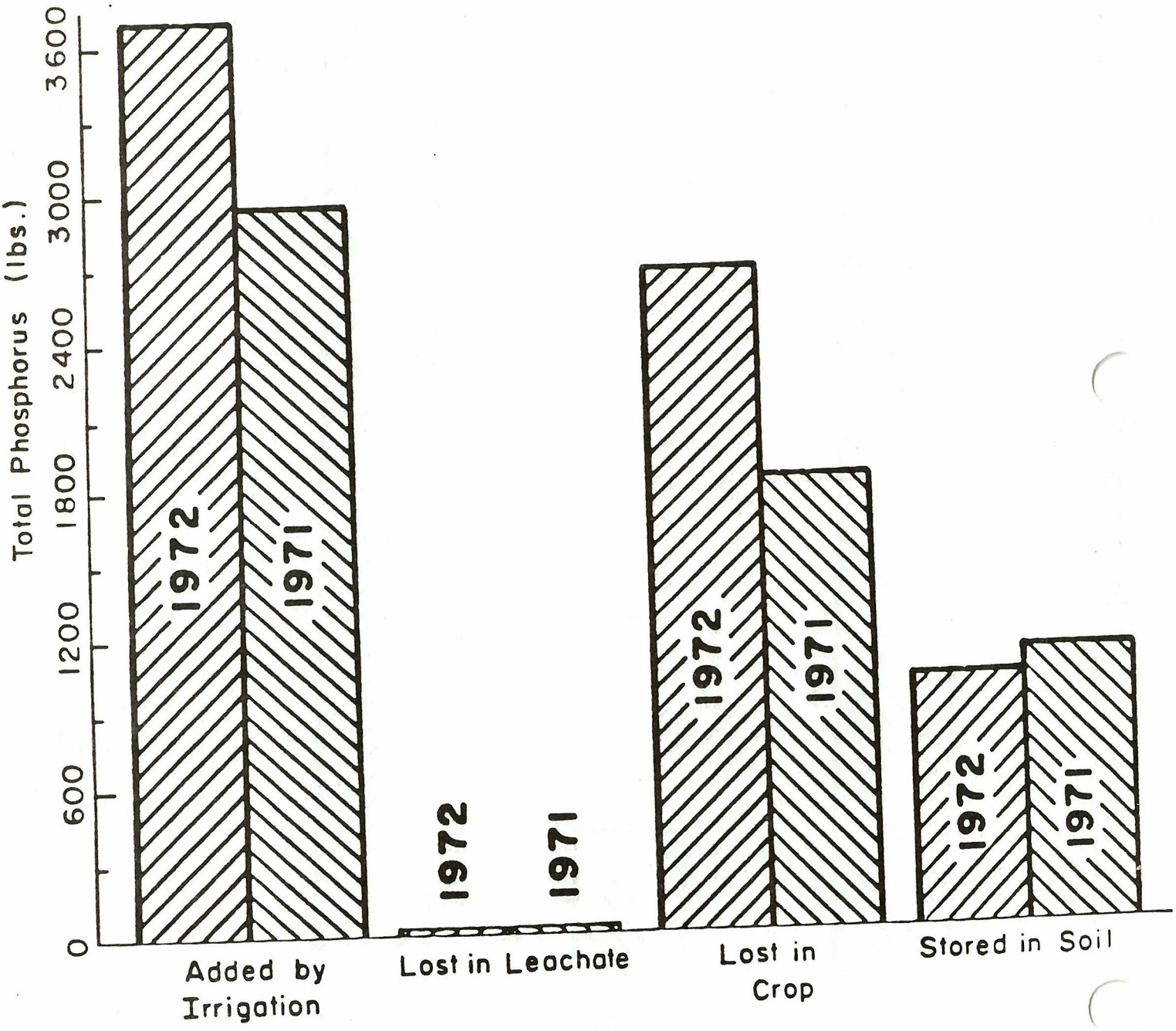


FIG. IV

1971 AND 1972 PHOSPHORUS BALANCE.

(Alfalfa Fields)



proceeded until October 25, and hence both water and nutrients were being added for a considerable period of time after all crop growth had ceased. If irrigation had been discontinued by about September 15, the seasonal water and phosphorus balance would have been much closer.

In areas where there is substantial precipitation during the non-growing season, it would appear reasonable that irrigation application could be cut back to just equal seasonal net evapotranspiration, because the off-season precipitation would be sufficient to leach any undesirably high concentration of impurities out of the root zone. However, in arid areas, it may be necessary to continue irrigation past the growing season to get some downward movement of unwanted impurities.

The design land area requirements will be dependent upon the design decision respecting the minimum amount of non-consumptive water that must be added in order to prevent impurity build-up in the root-zone soil.

MAXIMIZATION OF EFFLUENT APPLICATION: Where it is desired to minimize land requirements for effluent irrigation, it is imperative that a thorough understanding of all design parameters be achieved before design is finalized. The object is to plan an installation with as small an area as possible concomitant with maintaining acceptable conditions in the soil and in receiving waters for the intended life of the project.

The design situation is similar to that which occurs when the use of exfiltration ponds is being considered. In the case of irrigation, we are deliberately making use of the treatment capacity of the soil, although the crop does play a role in impurity removal. It is therefore apparent that a site which may be ruled inappropriate for the use of an exfiltration basin could be technically acceptable for a fairly high rate effluent irrigation project.

The best known example of this irrigation method is the Muskegon, Michigan project, in which some 27 x 10^6 US mgd (102,000 m^3/d) of combined municipal and pulp mill wastewater is being used to irrigate 5500 acres (2200 hectares) of corn[11]. Precipitation in the region averages 30 in/yr (76.2 cm/yr) with irrigation of aerated lagoon effluent adding up to 100 in/year (254 cm/yr) of applied water.

A system of underdrains with a total length of 65 miles (105 km) is used to collect the water in excess of crop requirements, and transport it to a surface water discharge.

Although the system has only been in operation since 1974, indications are that the present application rate of effluent can be maintained without significant leaching of macronutrients to the drainage system. However, this is only true because the concentrations of N and P in the lagoon effluent are about 4 mg/l and 1.5 mg/l respectively. A more conventional municipal effluent (with, say, 35 mg/l N and 8 mg/l P) would almost assuredly give rise to breakthrough conditions very quickly,as nutrients would be added to the site at much higher rates than the crop requirements dictate, and the sandy soil regions would allow the excess to pass into the drainage water very quickly (within a few years).

CONCLUSIONS

If it is desired to use land as a disposal site, or as a reuse process, for our wastes, and if it is desired to do so without causing health problems or unacceptable environmental damage, such projects should be approached cautiously. It is possible to build enough safety factors into regulations that land disposal, as regulated, could be used safely in every case. However, the costs of any such ventures would tend to make municipalities shy away from them in favour of more "conventional" solutions which may be more damaging to the environment.

If municipalities hope to convince senior governments not to over-regulate the use of land for wastewater disposal, they must be willing to conscientiously investigate the various parameters that affect the ability of a given piece of land to assimilate the impurities. The carrying-out of such site-specific studies should ensure that environmental standards are maintained if the project is built according to plan, and operated in a conscientious manner.

BIBLIOGRAPHY

1. Adriano, D.C., A.E. Erickson, A.R. Wolcott and B.G. Ellis, 1974 Certain Environmental Problems Associated with Long-term Disposal of Food Processing Wastes. In Processing and Management of Agricultural Waste, Cornell Agricultural Waste Conference, Ithaca, N.Y.

2. Bole, J.B., 1975 Wastewater Irrigation: Crop Suitability and Response. In Technology Transfer Seminar - Spray Irrigation of Treated Wastewater Effluents, Harrison Hot Springs, B.C.

3. Bryan, F.L. 1975. Diseases Transmitted by Foods Contaminated with Wastewater. Presented at the Conference on the Use of Wastewater in the Production of Food and Fiber, Oklahoma City, Oklahoma.

4. Oldham, W.K. and G.F. Kennedy, 1972. Influence of Septic Tank Effluent on Receiving Water Nutrient Gain from Groundwater. Report to the Canada - British Columbia Okanagan Basin Agreement, Victoria, B.C.

5. Oldham, W.K. 1973. Spray Irrigation of Sewage Treatment Plant Effluent at Vernon, B.C. Report No. 40, Canada - British Columbia Okanagan Basin Agreement, Victoria, B.C.

6. Oldham, W.K. 1973. Lysimeter Spray Irrigation Study at Penticton, B.C. Report to the Canada - British Columbia Okanagan Basin Agreement, Victoria, B.C.

7. Parizek, R.R., L.T. Kardos, W.E. Sopper, E.A. Myers, D.E. Davis, M.A. Farvell and J.B. Nesbitt, 1967. Wastewater Renovation and Conservation, Penn State Studies No. 23, Penn State University, University Park, Pennsylvania.

8. Parsons, D., C. Brownlee, D. Wetter, A. Maurer, E. Haughton, L. Kornder, and M. Slezak 1975. Health Aspects of Sewage Effluent Irrigation, Published by Pollution Control Branch, Gov't. of British Columbia Victoria, B.C.

9. Thomas, R.E., 1973, Spray - Runoff to Treat Raw Domestic Wastewater, In Proceedings, International Conference on Land for Waste Management, Ottawa, Ontario.

Ehlert, N., 1975, Spray Runoff Disposal of Waste Stabilization Pond Effluent, Project No. 71-1-19, Ontario Ministry of the Environment, July, 1975.

11. Walker, J.M., and Y.A. Demirjian, "Muskegon County, Michigan's Own Land Wastewater Treatment System", in State of Knowledge in Land Treatment of Wastewater, Vol. 1, U.S. Army Corps of Engineers Publication, pp. 417-428, Aug. 1978.

FUNDAMENTALS OF BIOLOGICAL PROCESSES FOR NITROGEN CONVERSION

by

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INTRODUCTION

Concern regarding nutrient removal can be traced back two decades when researchers correlated impairment of receiving water quality to discharge of nutrients from both municipal and industrial wastewaters. Two nutrients of primary concern are nitrogen and phosphorus. Control of these two nutrients is required to:

1) reduce the oxygen demand on the receiving water;
2) prevent eutrophication;
3) minimize human health hazards; and
4) render the effluents non-acutely lethal.

As phosphorus was identified as the major pollutant, intensive research programs conducted over the past decade have resulted in the development and full scale evaluation of phoshorus removal processes which are both practical and economical[1,2,3].

Control of nitrogen in wastewaters is much more complex and difficult than phosphorus control. Extensive full scale experience has resolved the choice between physical-chemical and biological processes for nitrogen control[4,5,6]. Biological systems have proven clear overall superiority.

Biological nitrogen removal is essentially a two-step process, nitrification followed by denitrification. In the nitrification step, under aerobic conditions, autotrophic nitrifying organisms oxidize ammonia to nitrate. In the denitrification step, nitrate is reduced to molecular nitrogen by heterotrophic organisms in the absence of molecular oxygen. The simplicity of the process requirements and the absence of waste by-products have encouraged research studies into biological nitrification/denitrification. In the United States and Canada, the majority of domestic and many industrial treatment plants employ biological treatment. These existing facilities can be used if upgrading for nitrogen conversion or removal is required.

The efficiency of biological nitrogen removal is correlated to the degree of nitrification achieved. The autotrophic nitrifiers

have lower yields and growth rates, and are more sensitive to environmental conditions than the heterotrophic denitrifiers. Consequently, nitrification probably will prove to be the crucial step in the nitrification/denitrification process.

BIOLOGICAL NITRIFICATION

Principles

Biological oxidation of inorganic nitrogen is almost entirely carried out in the aquatic environment by bacteria belonging to the family Nitrobacteraceae. The major well defined autotrophic bacteria which are known to oxidize ammonia to nitrite are the genera Nitrosomonas and Nitrosococcus. The genera Nitrobacter and Nitrocystis have been shown to oxidize nitrite to nitrate.

The autotrophic nitrifying organisms derive all their energy requirements for growth and cellular metabolism from the free energy released by the oxidation of an inorganic nitrogen substrate. The source of carbon used for growth is obtained from carbon dioxide or bicarbonate. The bacterial oxidation reactions of ammonium to nitrite, and nitrite to nitrate, can be represented by the following equations:

$$NH_4^+ + 1.5\ O_2 \longrightarrow 2\ H^+ + H_2O + NO_2^- \qquad (1)$$

$$NO_2^- + 0.5\ O_2 \longrightarrow NO_3^- \qquad (2)$$

As mentioned, these reactions provide the energy required for growth of the nitrifying organisms. The equations can be combined with equations for organism synthesis to form overall synthesis - oxidation equations by knowledge of the yield coefficients for the nitrifying organisms. Using reported values of actual cell yields, and assuming the empirical formulation of bacterial cells is $C_5H_7NO_2$, the following overall mass balances have been proposed[7]:

Nitrosomonas

$$55\ NH_4^+ + 5\ CO_2 + 76\ O_2 \longrightarrow C_5H_7O_2N + 54\ NO_2^- + 52\ H_2O + 109\ H^+ \qquad (3)$$

Nitrobacter

$$400\ NO_2^- + 5\ CO_2 + NH_4^+ + 195\ O_2 + 2\ H_2O \longrightarrow C_5H_7O_2N + 400\ NO_3^- + H^+ \quad (4)$$

On the basis of the equations, 20 mg of ammonium-nitrogen would produce only 3 mg of Nitrosomonas and approximately 0.5 mg of Nitrobacter. These yields are less than 10% of that normally observed for heterotrophic bacteria.

The growth of microorganisms may be expressed in terms of their doubling or generation time. Generation times of the autotrophic nitrifying bacteria are in the range of 10 to 30 hours[4]. Generation times for heterotrophic organisms, the bacteria responsible for BOD removal, are frequently reported as 20 to 40 minutes[8]. In a suspended growth or activated sludge system a direct consequence of the slow growth rate or long generation time of nitrifiers is the requirement of a sufficient solids retention time (SRT) or sludge age to retain an adequate population of these organisms. Solids retention time is a measure of the average retention time of the bacterial cells in the system. In an activated sludge system, the SRT is normally defined as the total mixed liquor suspended solids under aeration divided by the daily solids lost in the effluent or through sludge wasting.

Environmental Factors

Autotrophic nitrifying organisms are obligate aerobes. Numerous reports indicate that in order to ensure that dissolved oxygen (DO) is not a limiting factor in the nitrification reaction, a level not less than 2.0 mg/L must be maintained[9,10]. At lower DO levels the nitrification rate will be significantly reduced (Figure 1). The stoichiometric oxygen requirements based on Equations 6 and 7 are 3.45 mg oxygen/mg ammonium-nitrogen and 1.14 mg oxygen/mg nitrite-nitrogen.

Nitrification, like most bacterial processes, is affected by pH conditions. Generally, optimum conditions have been found to exist between 8.0 and 9.0 (Figure 2). Variations in pH optima could be due to shock effects in adjusting culture conditions or improper

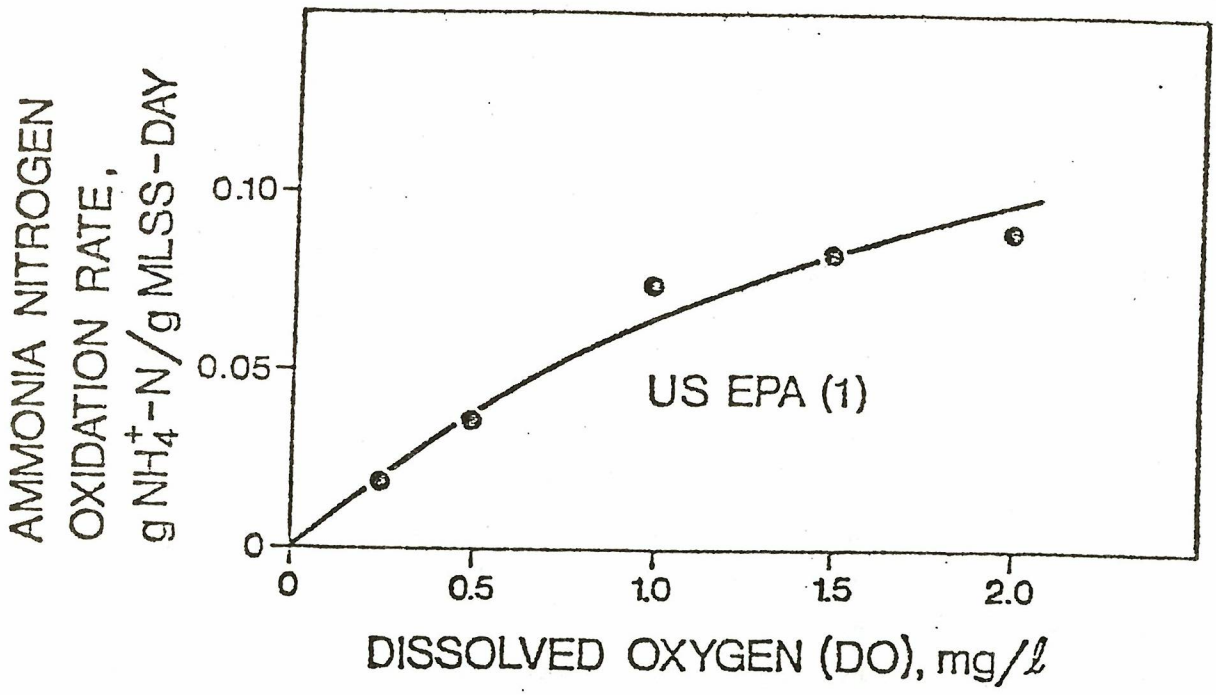


FIGURE 1. DEPENDENCY OF NITRIFICATION RATE ON DISSOLVED OXYGEN CONCENTRATION

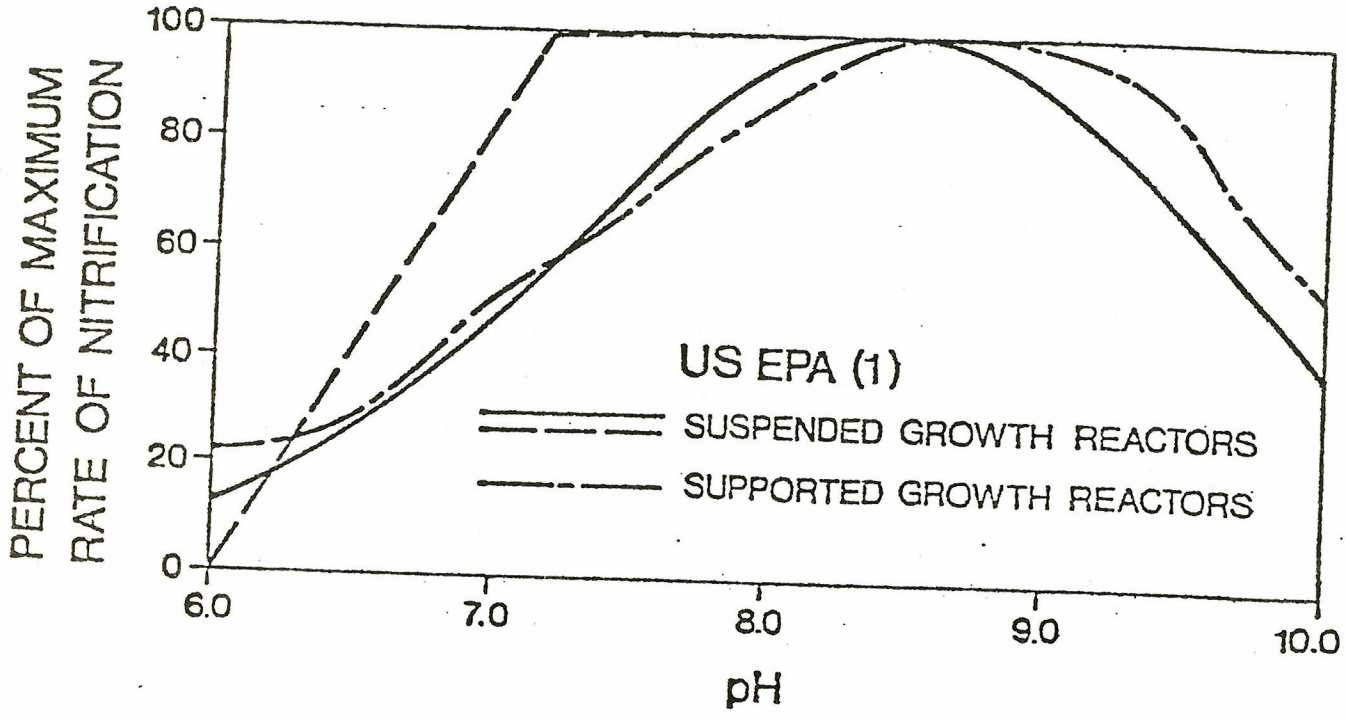


FIGURE 2. DEPENDENCY OF NITRIFICATION ON pH

acclimation. Haug and McCarty [7], using a submerged aerobic filter, found that after an acclimation period of approximately 10 days, the rate of nitrification at pH 6.0 was similar to rates within the normal range of operation, a pH of 7 to 8.5.

Nitrificaiton is a hydrogen ion producing reaction (Equation 1). Since pH values less than 6.0 would likely affect the nitrification rate, the alkalinity of the waste is an important consideration. Assuming the pH is less than 8.5, the hydrogen ions produced during nitrification react with the bicarbonate in the wastewater, resulting in an increase in CO_2 concentration and a decrease in bicarbonate alkalinity according to:

$$H^+ + HCO_3^- \longrightarrow CO_2 + H_2O \qquad (5)$$

Based upon Equations 1 and 5, approximately 7 mg of bicarbonate alkalinity, expressed as $CaCO_3$, are required to neutralize the hydrogen ions produced during the oxidation of 1 mg of ammonia-nitrogen. Calculations involving carbonic acid equilibria show that for wastewater with an alkalinity of 200 mg/L as $CaCO_3$, approximately 20 mg/L ammonia-N could be oxidized before the pH dropped below 6.0 if all the CO_2 produced remained in solution[7]. In most nitrifying reactors, the CO_2 is stripped from solution tending to help maintain a neutral pH. In operating a rotating biological contactor for carbon (BOD) removal-nitrification of a municipal sewage with moderate alkalinity (approximately 120 mg/L as $CaCO_3$), Wilson[11] found that 15 to 20 mg/L ammonia-N were nitrified and the pH was never less than 6.8. Sutton[12] observed similar results in operating suspended growth carbon removal-nitrification systems.

The process of nitrification, occurring over a range of approxiamtely 4 to 45°C with optima at about 35°C for Nitrosomonas[13] and 35 to 42°C for Nitrobacter[14,15], has been shown to be a strong function of temperature. In a suspended growth nitrification system differences in the reported temperature sensitivity may be due to differences in reactor SRT (Figure 3). Increasing the SRT results in a decrease in the temperature sensitivity. Supported growth systems (trickling

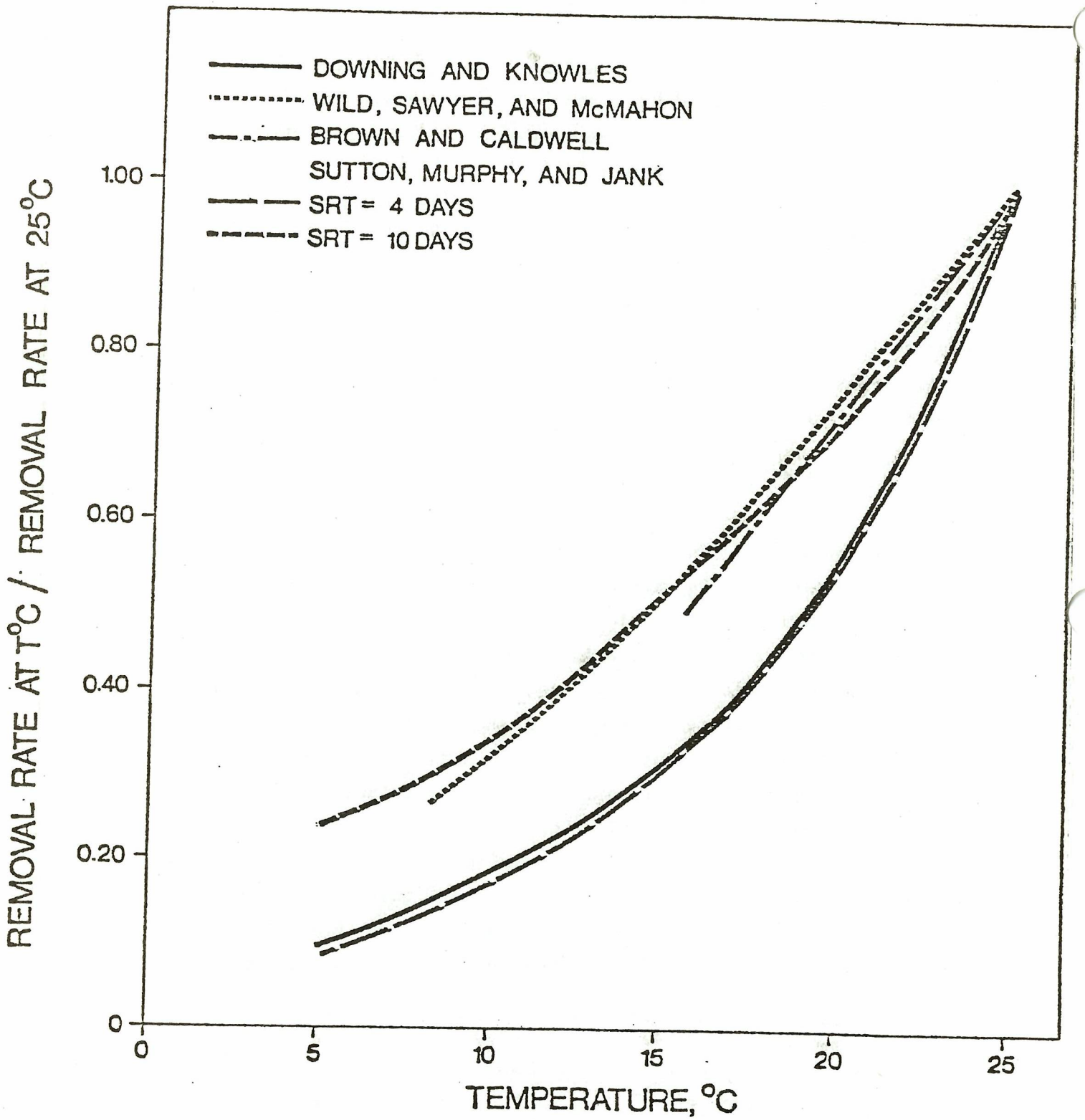


FIGURE 3. DEPENDENCY OF NITRIFICATION RATE ON TEMPERATURE

filter, rotating biological contactor, etc.), which could be expected to be operating at high SRT's, show reduced temperature sensitivity for nitrification[16,17].

Nitrifying organisms, especially Nitrosomonas, are susceptible to a number of inhibitors which may be present in municipal and industrial wastewaters. A number of metals are toxic to nitrifiers but the concentration required to cause inhibition is dependent on the state of the culture. Copper, silver, mercury, nickel, chromium, and zinc all may inhibit nitrification under certain conditions. Downing, Tomlinson, and Truesdale[18] identified a large number of organic materials which caused a reduction in nitrification rates in suspended growth systems. Thiourea, phenol, phenolic compounds, and halogenated solvents were some of the potential inhibitors identified, although nitrifiers can acclimate to phenolic compounds. In screening the organic materials as nitrification inhibitors, unacclimated organisms were used and, therefore, the results may not reflect what actually would occur in a treatment plant. Hockenbury and Grady[19] tested 52 industrially important organic compounds for their effects on Nitrosomonas and Nitrobacter. Twenty compounds were found to inhibit ammonia oxidation when present in the medium at a concentration of 100 mg/L. Three were found to inhibit nitrite oxidation. The three strongest inhibitors of ammonia oxidation were dodecylamine, aniline, and n-methylaniline. Anthonisen[20] recently has identified levels of free ammonia (NH_3) and free nitrous acid (HNO_2) at which inhioition of Nitrosomonas and Nitrobacter occurs. The inhibitory concentrations in the medium ranged from 0.1 to 150 mg/L for NH_3 and 0.2 to 2.8 mg/L for HNO_2. The concentrations of NH_3 and HNO_2 are dependent on the concentrations of total ammonia and total nitrite, respectively, and the pH and temperature conditions.

In the design of nitrification systems for industrial waste treatment, the possibility of toxic inhibition must be recognized and unless source control is implemented, spill basins or toxicity removal processes may be required.

Reactor Kinetics and Design

Process design requires rate data to be expressed as parameters which are useful from an engineering point of view. In order to describe the substrate removal rate in any biological waste treatment process, an overall kinetic expression compatible with the fundamentals of microbiology, kinetics, system hydraulics, and transport phenomena, is necessary.

In the nitrification process, the growth of Nitrosomonas and Nitrobacter are limited by the concentration of ammonia and nitrite, respectively. Since the maximum growth rate of Nitrobacter is greater than the maximum growth rate of Nitrosomonas, the rate limiting step in the nitrification reaction is the conversion of ammonia to nitrite. The growth rate of Nitrosomonas (NS) under steady-state conditions, is normally[4] described by:

$$U_N = U_N^* \frac{N}{K_N + N} \tag{6}$$

where: U_N = growth rate of NS ($days^{-1}$);
U_N^* = maximum growth rate of NS ($days^{-1}$);
N = NH_4^+-N concentration; and
K_N = N concentration at which $U_N = 0.5\ U_N^*$ (mg/L)

The oxidation of ammonia can be related to the NS growth rate according to:

$$\frac{1}{X_N}\frac{dN}{dt} = \frac{U_N}{Y_N} = q_N = q_N^* \frac{N}{K_N + N} \tag{7}$$

where: X_N = NS mixed-liquor volatile suspended solids concentration (mg/L);
q_N^* = ammonia oxidation rate (mg NH_4^+-N oxidized/mg NS VSS·d);
q_N = maximum ammonia oxidation rate (mg NH_4^+-N oxidized/mg NS VSS·d);
K_N = N concentration at which $q_N = 0.5\ q_N^*$ (mg/L); and
Y_N = organism net yield coefficient (mg NS VSS grown/mg NH_4^+-N removed.

Equation 7 indicates a zero-order relationship between the oxidation rate and substrate concentration when the value of K_N is much less than N. The zero-order nature of the nitrification reaction down to low ammonia-nitrogen concentrations (1 to 3 mg/L) has been reported by numerous authors in the treatment of municipal wastewaters[21,22,23]. Zero-order relationships have also been found in the treatment of organic chemical[24], petrochemical[25], textile[25] and agricultural[26] wastewaters.

The SRT and the growth rate of all organisms (U) in an activated sludge system are related by:

$$SRT = \frac{1}{U} \tag{8}$$

Consequently, the growth rate of Nitrosomonas is fixed according to the SRT of the system. Since the ammonia oxidation rate (q_N) is a function of the growth rate of NS (Equation 8), then q_N is fixed according to the SRT. The relationship between SRT and the growth rate or organisms implies a procedure for design and control of a nitrification system. Having determined the SRT necessary to establish nitrification, the relatonship between growth rate and substrate removal can be used to establish the other design parameters (hydraulic retention time, mixed liquor volatile suspended solids, etc). In an activated sludge system in which a high degree of BOD removal and nitrification are required, the design procedure involves using the growth rate and substrate removal rate relationship for the heterotrophic population to establish the other design parameters[4,8].

The relationship describing the oxidation of ammonia (Equation 8) reduces a constant (q_N^*) if the nitrification reaction is considered to be zero-order, according to:

$$\frac{1}{X_N}\frac{dN}{dt} = q_N^* \tag{9}$$

Equation 9 representes a simplified relationship for design provided X_N can be determined. In aerobic nitrification systems in which carbon oxidation is occurring together with nitrification, the

fraction of X_N in the total volatile suspended solids (X) will depend on the wastewater BOD to ammonia ratio. To date no rapid and simple technique exists for the determination of the fraction X_N in X. In the treatment of municipal wastewaters, wide variatons in the values for the maximum ammonia oxidation rate based on the total volatile suspended solids q*), under similar environmental conditions (Temperature, pH, etc.), can be related to differences in the wastewater BOD to ammonia ratio[4,27]. The good agreement between reported SRT values necessary for nitrification of municipal wastewaters[28,29] can be related to the independence of SRT and the wastewater BOD to ammonia ratio. Consequently, the SRT design technique represents a more universal procedure for design of municipal systems.

In the design of nitrification systems for industrial waste treatment, the wastewater characteristics will vary from plant to plant. Therefore, pilot plant or bench scale studies will normally be required to establish the SRT necessary for nitrification or to determine the maximum ammonia oxidation rate (q*). The use of an ammonia oxidation rate for design has also been proposed by Eckenfelder[30]. This procedure, although strictly an empirical approach and subject to various limitations[31], has been found to be an effective description of nitrification results.

BIOLOGICAL DENITRIFICATION

Principles

Biological denitrification is achieved under anoxic (absence of molecular oxygen) conditions by heterotrophic micoorganisms that utilize nitrate as a hydrogen acceptor when an organic energy source is available. A wide variety of common facultative bacteria such as Pseudomonas, Micrococcus, Achromobacter, Spirillum and Bacillus have been reported to accomplish denitrification[32].

The reduction of nitrate occurs through a series of complicated enzyme catalyzed reactions which can follow either the assimilatory or dissimilatory route. In assimilatory denitrification, nitrate is reduced to ammonia that is used in cell synthesis. In dissimilatory denitrification, in the absence of molecular oxygen, nitrate

serves as the hydrogen acceptor in the oxidation of the carbon substrate, producing energy for cell growth. The gaseous end products of the reactions are principally dinitrogen (N_2).

In oxygen respiration, the same series of reversible enzymatic reactions are involved in transferring electrons from organic substrates to molecular oxygen. Only a different terminal enzyme is required for nitrate respiration and this is formed by the bacteria in the absence of oxygen. Consequently, nitratebacteria dissimilation is inhibited by oxygen. The occurrence of denitrification under aerobic conditions, as reported by some researchers[33,34] is most likely due to a dissolved oxygen gradient across the bacterial floc.

In a nitrogen removal system treating municipal or industrial wastewater containing carbonaceous material and organic nitrogen, or inorganic nitrogen in the form of ammonia but little or no oxidized nitrogen, the system specified will often be a post-denitrification system. In this case, denitrification will follow the processes of carbon oxidation and nitrification. Therefore, the residual organic carbon from the aerobic processes will not be of sufficient concentration, not in a readily useable form for use by the denitrifying population as an organic carbon and energy source. The carbon and energy source may be provided internally by by-passing part of the raw sewage to the denitrification reactor or by the endogenous respiration of the sludge, or externally by the addition of a carbon and energy source. The use of raw sewage has an obvious economic advantage but results in ammonia appearing in the effluent from the denitrification system. Denitrification utilizing the endogenous respiration of sludge is feasible but the reaction rates obtained are much lower than the rates found with systems utilizing external carbon and energy sources[4]. Various organic materials such as sugar, acetic acid, and acetone, have been found satisfactory as external electron donors but methanol appears to be the optimum choice[35]. In an industrial environment, carbon-rich wastes, if available, are attractive alternatives[36].

In order to write stoichiometric equations for the denitrification reaction, the composition of the carbon source must be known.

Considering denitrification to be a two-step process with methanol as the carbon and energy source, the following reactions can be written:

Step 1: $$3\ NO_3^- + CH_3OH \longrightarrow 3\ NO_2^- + CO_2 + 2\ H_2O \quad (10)$$

Step 2: $$2\ NO_2^- + CH_3OH \longrightarrow N_2 + CO_2 + H_2O + 2\ OH^- \quad (11)$$

Overall: $$6\ NO_3^- + 5\ CH_3OH \longrightarrow 3\ N_2 + 5\ CO_2 + 7\ H_2O + 6\ OH^- \quad (12)$$

In addition to the methanol required for nitrate dissimilation (Equation 10), an amount is necessary for bacterial growth. An overall mass balance which consideres this requirement is as follows[37]:

$$CH_3OH + 0.92\ NO_3^- + 0.92\ H^+ \longrightarrow 0.06\ C_5H_7O_2N + 0.43\ N_2 + 0.7\ CO_2 + 2.25\ H_2O \quad (13)$$

On the basis of this equation, 20 mg of NO_3-N would require 19 mg of CH_3OH-C and would produce 10.5 mg of bacteria. Denitrification studies have determined that the actual methanol requirement and biomass production can be estimated from the following relationship[35]:

Methanol Requirement: $$C_m = 2.47\ N_o + 1.53\ N_1 + 0.87\ D_o \quad (14)$$

Biomass Production: $$C_b = 0.53\ N_o + 0.32\ N_1 + 0.19\ D_o \quad (15)$$

where: C_m = required methanol (mg/L);
C_b = biomass production (mg/L);
N_o = initial NO_3-N concentration (mg/L);
N_1 = initial NO_2-N concentration; and
D_o = the dissolved oxygen concentration in the wastewater entering the system (mg/L).

A number of researchers[4] have obtained experimental results which verify that these expressions provide reasonable estimates of methanol requirements and biomass production.

Environmental Factors

Reduced denitrification rates in the presence of dissolved oxygen can be accounted for by the effect of oxygen on the formation of necessary enzymes, or activity of preformed enzymes, required for dissimilatory denitrification[38]. Although denitrification will be observed in systems at low oxygen tension, a compeltely anoxic reactor will maximize nitrate reduction.

The optimum pH for denitrification is usually reported at conditions near neutrality (Figure 4). Nitrate dissimilation is a basic reaction (Equation 13) with hydrogen ion consumption tending to raise the pH. When used together with nitrification, the nitrification/denitrification sequence tends to maintain a neutral pH[42,43].

Denitrification is reported to occur over a temperature range from 0 to 50°C with an optimum approaching 40°C[41]. There is reasonably good agreement between authors[4] concerning the degree of temperature sensitivity over the range normally encountered in municipal wastewater treatment (Figure 5). An extensive list of values for various coefficients normally used to express the reaction rate and temperature dependency in biological systems has been presented elsewhere[44].

In biological nitrification/denitrification the autotrophic nitrifiers are more sensitive to environmental conditions than the heterotrophic denitrifiers. Ensuring that there are no toxicants in the wastewater, limiting the nitrification rate will also ensure that the denitrification process will be unaffected.

Reactor Kinetics and Design

In denitrification, carbon and nitrate are the important substrates involved in the biological reaction. The denitrification process has been applied to the treatment of wastewaters at nitrate concentrations of 10 000 mg/L[46]. Numerous researchers have shown that the denitrification rate is independent of nitrate concentration (zero-order) down to values of less than 1.0 mg/L NO_3-N in suspended growth reactors[45,47]. In supported growth systems, both zero-order[44,48]

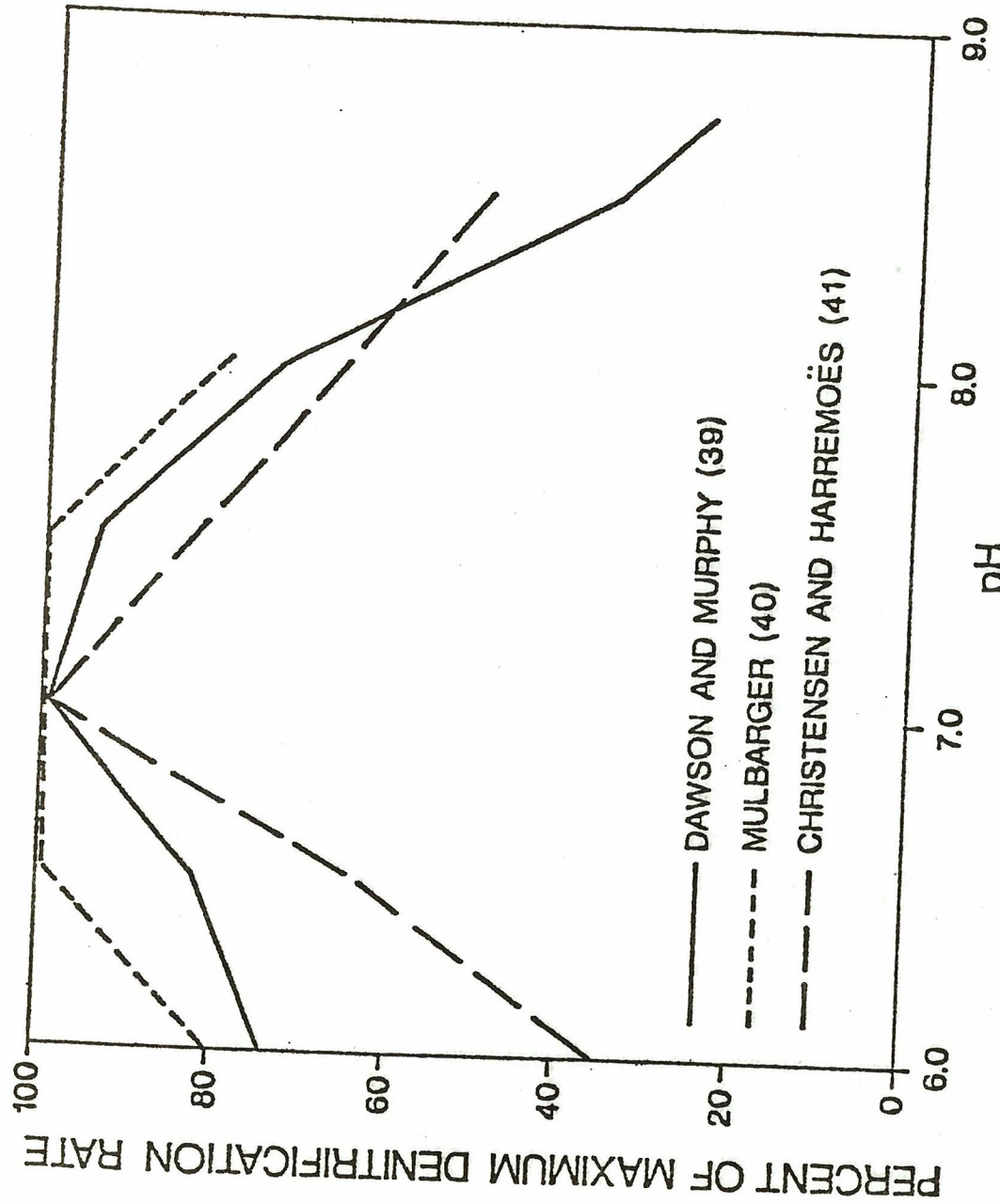


FIGURE 4. DEPENDENCY OF DENITRIFICATION ON pH

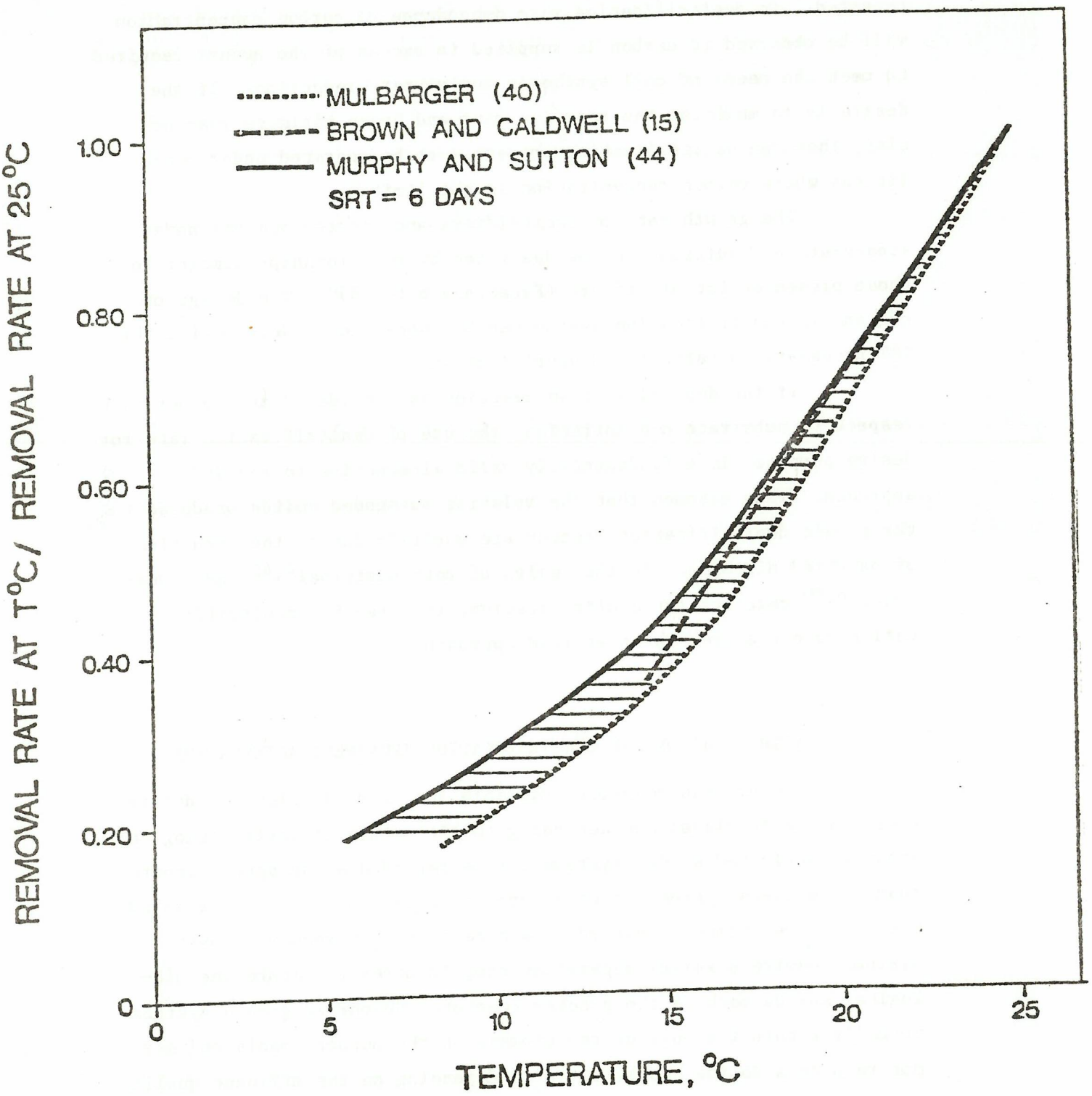


FIGURE 5. DEPENDENCY OF DENITRIFICATION RATE ON TEMPERATURE

and half-order[47] nitrate concentration dependencies have been reported. No denitrification rate dependence on carbon concentration will be observed if carbon is supplied in excess of the amount required to meet the needs of cell synthesis and nitrate reduction. If the desire is to maximize the reaction rate and hence minimize reactor size, then the denitrification process must be operated under conditions where carbon concentration is not limiting.

The growth rate of denitrifiers and nitrate removal under steady-state conditions can be described by relationships similar to those presented for nitrifiers (Equations 6 to 18)[4]. The design of the anoxic denitrification system can be specified by determining the SRT necessary to establish denitrification.

If the denitrification reaction is considered zero-order with respect to substrate concentration, the use of denitrification rate for design purposes is a fundamentally valid alternative to the SRT approach. This assumes that the volatile suspended solids produced in the anoxic denitrification reactor are entirely due to the reduction of oxidized nitrogen. In the design of both municipal[4,42] and industrial[46,50] reactors for denitrification, the use of a denitrification rate has been the preferred approach.

NITRIFICATION AND DENITRIFICATION TREATMENT ALTERNATIVES

The process reactors available for nitrification and denitrification can be classified according to the nature of their biological growth. Activated sludge systems can be regarded as suspended growth reactors whereas systems in which growth occurs on, or within a solid media, can be termed supported growth reactors. Suspended growth systems require a solids separation step in order to return the biological solids back to the process reactor. Supported growth systems normally retain the bulk of the biomass on the support media and may not require a solids separation step depending on the effluent quality requirements.

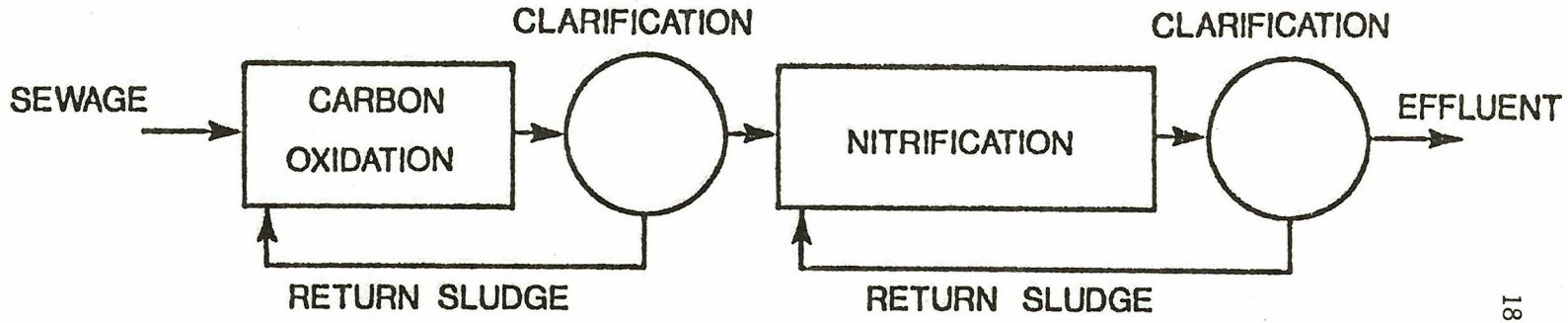
Suspended Growth Nitrification/Denitrification Alternatives

The two basic process schemes available for carbon oxidation and nitrification in the activated sludge process are the combined sludge process and the two-stage separate sludge process (Figure 6). In the combined sludge process, carbon oxidation and nitrification are carried out using the same sludge. In the separate sludge process, the biological reactions are carried out by different microorganisms in separate reactors. Other combined sludge systems which can be designed to incorporate partial or complete nitrification include: contact stabilization, extended aeration, step aeration and oxygen activated sludge plants. Combined sludge systems for carbon removal and nitrification of municipal and industrial wastewaters have proven to be a viable alternative to the separate sludge system[51,52,53,54]. In the treatment of certain industrial wastes, the protection of the sensitive nitrifying organisms in a separate reactor may be necessary in order to achieve high degrees of nitrification[55,56,57].

A number of combined sludge systems can be designed to incorporate both carbon oxidation, nitrification, and denitrification (Figure 7). High degrees of total nitrogen removal have been reported using these systems for the treatment of both municipal[41,43,58] and industrial[23,50,59] wastes. The pre-denitrification/nitrification system (Figure 7) is particularly suitable for the treatment of industrial wastes containing high concentrations of degradable organic carbon, oxidized nitrogen, and ammonia. The organics in the wastewater provide the carbon substrate for nitrate reduction in the first stage. In the second stage any remaining organics are oxidized and the ammonia converted to nitrate. If further nitrate reduction is required, an additional denitrification stage may be added creating an alternating denitrification/nitrification system[42] (Figure 8).

The oxidation ditch has received attention as a combined sludge nitrogen removal process[59]. The aeration rotors (Figure 8) induce cycling of the mixed liquor creating aerobic and anoxic zones and consequently promoting nitrification and denitrification. To date, the most popular nitrogen removal system in the United States[4] involves separate or combined sludge carbon removal and nitrification followed by

TWO-STAGE SEPARATE SLUDGE



COMBINED SLUDGE

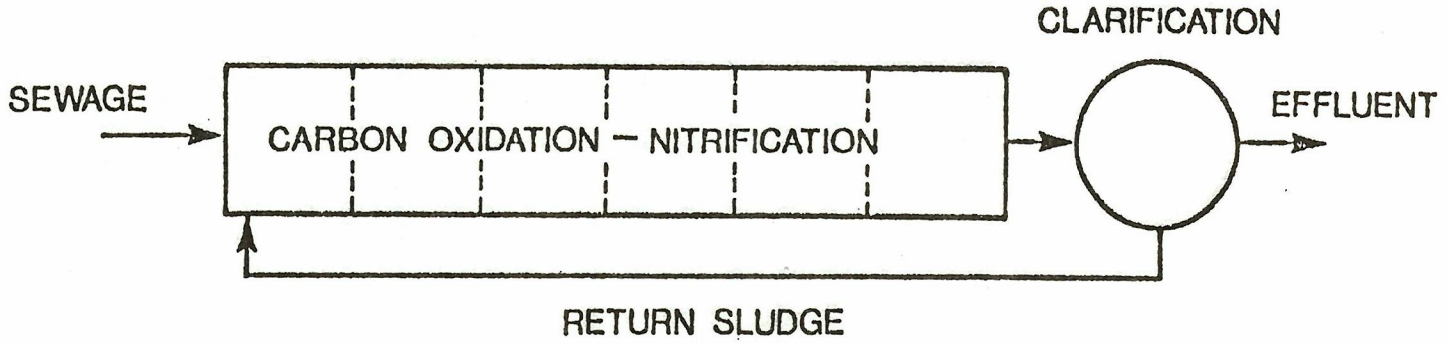
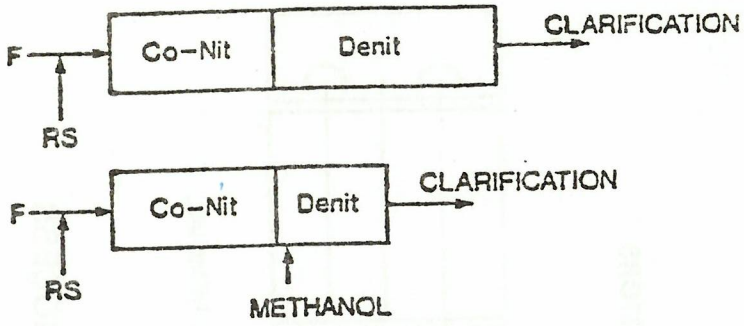


FIGURE 6. SEPARATE AND COMBINED SLUDGE CARBON OXIDATION AND NITRIFICATION SYSTEMS

CARBON OXIDATION - NITRIFICATION POST - DENITRIFICATION

NO EXTERNAL CARBON SOURCE ADDITION

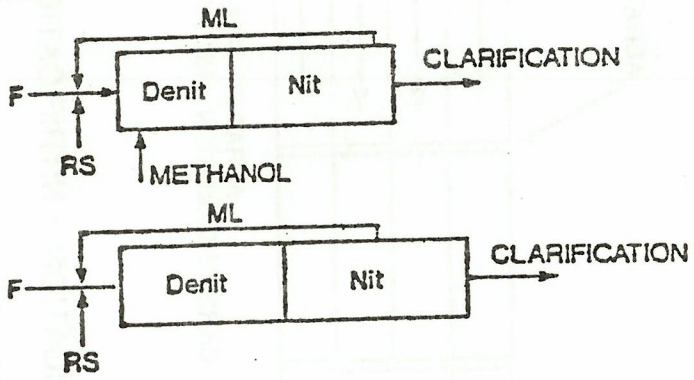
EXTERNAL CARBON SOURCE ADDITION



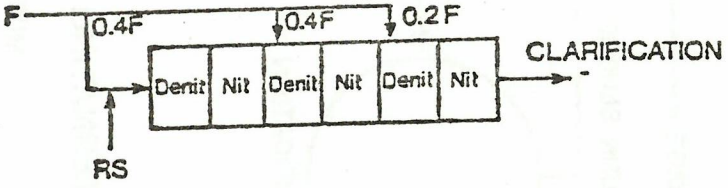
PRE - DENITRIFICATION NITRIFICATION

EXTERNAL CARBON SOURCE ADDITION

NO EXTERNAL CARBON SOURCE ADDITION



STEP - FEED NITRIFICATION - DENITRIFICATION



Note : Co refers to carbon oxidation, Nit to nitrification, Denit to denitrification, F to feed, RS to return sludge, and ML to mixed - liquor

FIGURE 7. COMBINED SLUDGE SYSTEMS FOR CARBON OXIDATION NITRIFICATION AND DENITRIFICATION

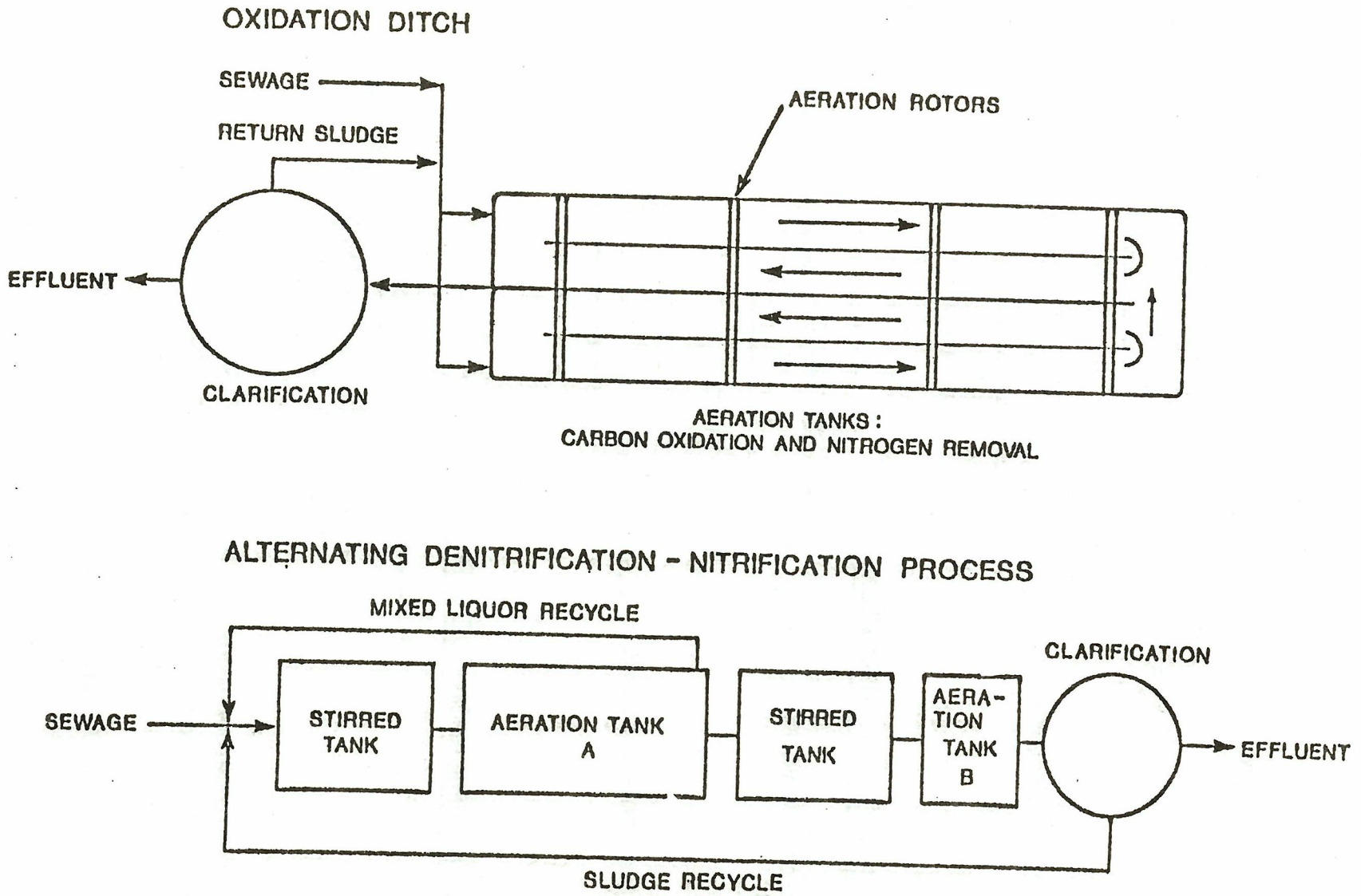


FIGURE 8. COMBINED SLUDGE NITROGEN REMOVAL SYSTEMS

separate sludge denitrification with the addition of an external carbon source (Figure 9). However, recent price increases for methanol have made this process uneconomical.

Supported Growth Nitrification/Denitrification Alternatives

A limited number of supported growth systems have been investigated for nitrogen control.

Trickling filters, submerged or expanded bed aerobic filters, and rotating biological contactors (Figure 10) have been used for carbon removal and/or nitrification.

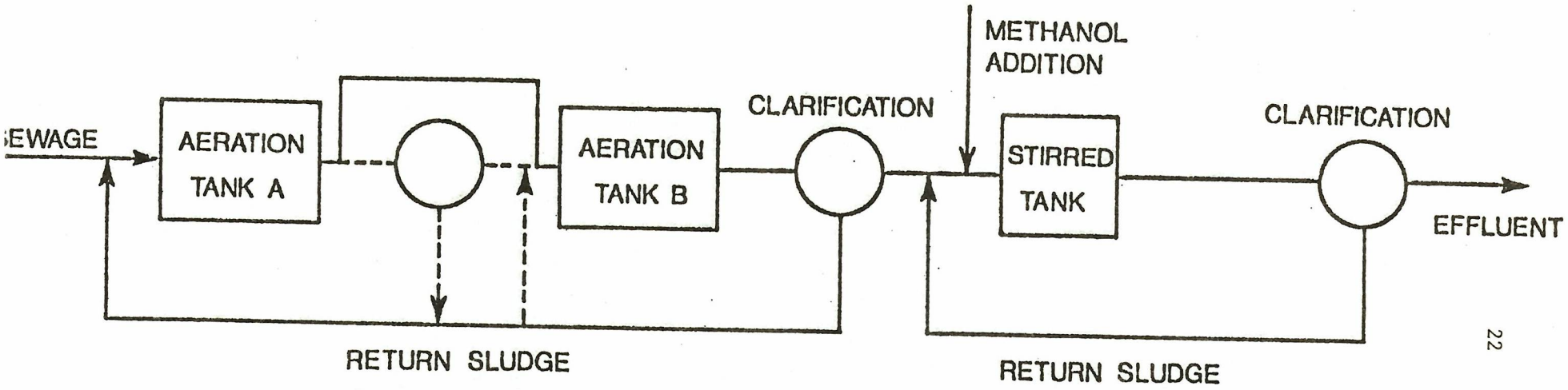
Rocks, wood slats, and plastic sheets have all been used as media for trickling filters. In submerged upflow packed bed reactors, stones, gravel, plastic, and other types of media can be employed. Although these packed tower reactors can be designed economically for nitrification down to ammonia-nitrogen levels approaching 2.0 mg/L, removal below this is difficult to attain[4].

Expanded or fluidized bed systems are currently in the development stage. In these systems, biological growth occurs on a media such as sand which is fluidized by the incoming wastewater. The results reported to date indicate that very high substrate removal rates are attainable[60].

The rotating biological contactor consists of a series of closely spaced discs anchored to a shaft and partially submerged in the wastewater. The rotation of the shaft alternately expose the biological slime growing on the disc surfaces to the waste and then to the atmosphere. Under proper loading conditions, successful carbon removal-nitrification can be obtained[11,61].

A number of reactors consisting of columns containing different types of packed media are available for biological denitrification (Figure 11). The column systems generally fall into one of the following categories:

1) Fixed bed upflow reactors with porous media of varying size exhibiting minor pressure gradients.
2) Expanded bed upflow reactors exhibiting a constant pressure gradient.
3) Fixed bed downflow or upflow reactors exhibiting an increasing pressure gradient necessitating backwashing or flushing.



AERATION TANK A AND B: CARBON OXIDATION AND NITRIFICATION

STIRRED TANK : NITRATE REMOVAL

FIGURE 9. COMBINED OR SEPARATE SLUDGE CARBON OXIDATION AND NITRIFICATION AND SEPARATE SLUDGE DENITRIFICATION

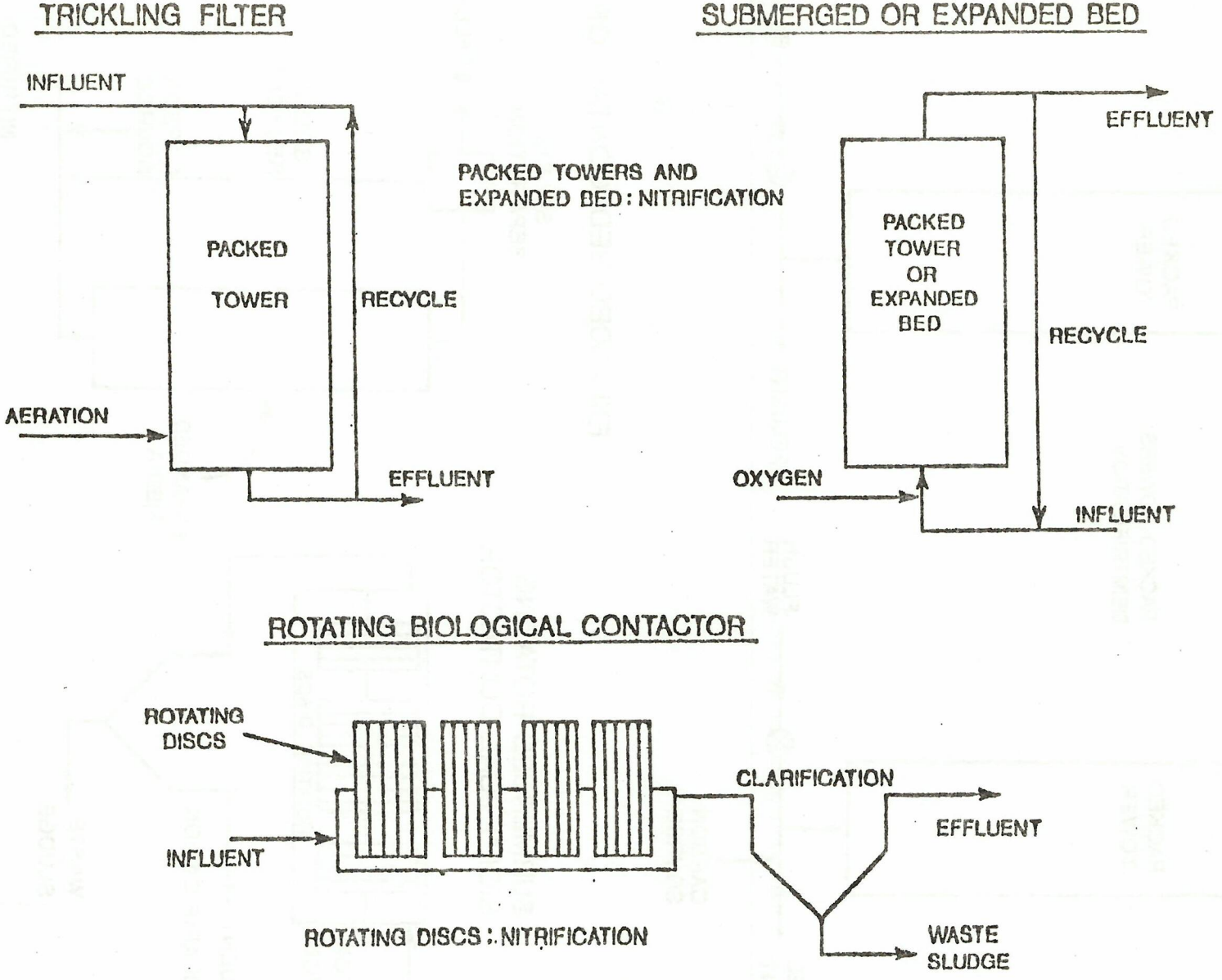


FIGURE 10. SUPPORTED GROWTH NITRIFICATION SYSTEMS

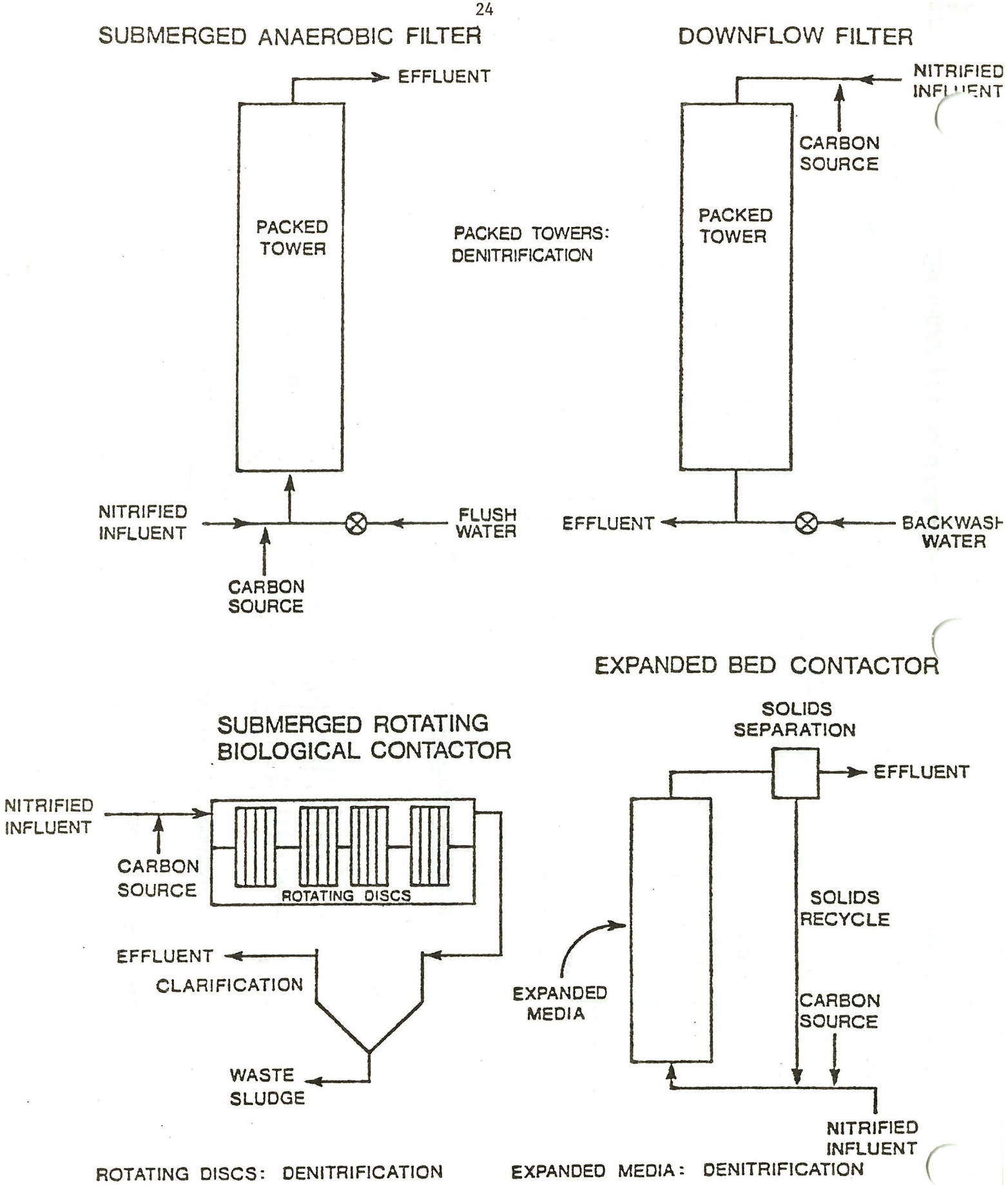


FIGURE 11. SUPPORTED GROWTH DENITRIFICATION SYSTEMS

The column reactors of the first category have shown significant but variable nitrate removal[11]. The expanded bed and pressure filters appear tobe acceptable supported growth systems for denitrification based on limited design information[60,63].

The rotating biological contactor can be employed for denitrification (Figure 11) by submerging the rotating discs below the liquid level to minimize oxygen transfer and maintain anaerobic conditions within the biological film. Preliminary results indicate that significant nitrate removals are possible[11].

The large number of suspended and supported growth systems complicates the task of process selection and design. Certain fundamental considerations must be made in applying a particular process to a certain situation. These include economic feasibility, degree of treatment required, type of ultimate disposal problem that may result, compatibility of other treatment processes, and operational problems. Any comparative assessment of alternative nitrification-denitrification schemes does not have general validity as the specifics of individual circumstances will determine the optimum sequence.

PRACTICAL DESIGN AND OPERATING CONSIDERATIONS

In the design of biological nitrogen control facilities it must be remembered that nitrification follows zero-order kinetics to substrate levels of $\sim$0.5 mg $NH_3 \cdot L^{-1}$ and denitrification to $\sim$1 mg NO_3-N$\cdot L^{-1}$. Thus, full scale systems should not be expected to produce an effluent quality of less than 2 mg TN$\cdot L^{-1}$. This is true only if no residual refractory organic nitrogen is present. In domestic wastewater, the refractory organic nitrogen is normally 1 to 2 mg$\cdot L^{-1}$ and in industrial wastewaters the value will be much higher.

Nitrification will normally be the rate-limiting step and is, therefore, the critical design parameter. The preferred design procedure is the SRT approach. This procedure eliminates the confounding effect of variable C/N feed ratio, as the minimum SRT required for nitrification is independent of C/N ratio. Design of facilities

treating exclusively municipal wastewater can be based on existing information. However, when treating wastewaters containing industrial discharges, bench scale studies are required to determine the minimum SRT requirements. In the treatment of full strength industrial wastewaters, the presence of inhibitory materials is likely and consequently much higher SRT's are normally required to achieve nitrification. From a theoretical viewpoint, SRT and HRT are independent. However, there are certain physical and other limitations. As the HRT is reduced, the mixed liquor suspended solids increase, and since oxidation of carbonaceous material follows first-order reaction kinetics, FOC in the effluent increases. Thus, from a practical point of view, there is some minimum HRT based on the desired MLSS level, and in the treatment of industrial wastes, based on required effluent FOC value. Furthermore, in the treatment of industrial wastewaters, a reduction in HRT could increase the level of some "inhibitory" compound in the effluent and thus alter the nitrification kinetics.

A parameter of major significance is feed variability. Once again, this is specifically true in the design of industrial nitrification systems. To overcome the effect of variability, the design SRT must be greater than the minimum SRT required to maintain nitrification under the appropriate environmental conditions. As a general rule of thumb, the following applies[4]:

$$\frac{SRT_{design}}{SRT_{min}} > \frac{TKN_{max}}{TKN_{mean}}$$

Practical experience treating both municipal and industrial wastewaters has indicated consistent nitrification can be achieved provided the ratio TKN_{max}/TKN_{mean} is equal to or less than 2.0[5,63,64].

Total nitrogen removal systems can comprise separate or combined sludge systems, operating both in the pre or post denitrification mode. Separate sludge systems have higher unit removal rates and consequently require lower reactor volumes. This advantage is normally

overshadowed by the advantages attributed to combined sludge systems, namely:

1) Reduction of clarification facilities.
2) Improved sludge properties.
3) Reduced alkalinity or acidity requirements.
4) Reduced aeration requirements.
5) Reduced carbon requirements for DN (predenitrification mode).

Single sludge systems for combined carbon oxidation, nitrification and denitrification appear to be the most cost-effective system for nitrogen control[5,50,64,65]. Recent studies have indicated that the anoxic or aerobic periods created in these flow sheets do not affect nitrification kinetics[63]. Single sludge systems can be operated in the predenitrification or post denitrification mode. Sutton[63] demonstrated no difference in denitrification performance of these two systems (Figure 12). The advantage of the predenitrification mode is that the organics present in the wastewater can serve as the electron donor and energy source for the denitrification reaction, thus eliminating or substantially reducing the cost for an external carbon source. Research has also indicated that raw sewage (Figure 13) and many industrial wastewaters (Table 1) provide suitable carbon sources for denitrification[36,63].

A disadvantage of the predenitrification system is the higher carbon requirement exhibited[63] (Figure 14). In this study a total of 8 mg filterable COD was required per mg of NO_T-N reduced compared to ~4 to 5 mg required for the post DN system, which is roughly the stoichiometric requirement. For the predenitrification system in this study, the feed COD/TKN ratio had to be in excess of 13 to ensure complete denitrification (Figure 15). This compares favourably with the results treating coke plant wastewaters[64] (Figure 16). A possible explanation for the higher carbon requirements for the pre-denitrification mode is the fact that this mode tends to operate with a higher C/N ratio, and it has been shown the C/N ratio can affect substrate consumption[36] (Figure 17.)

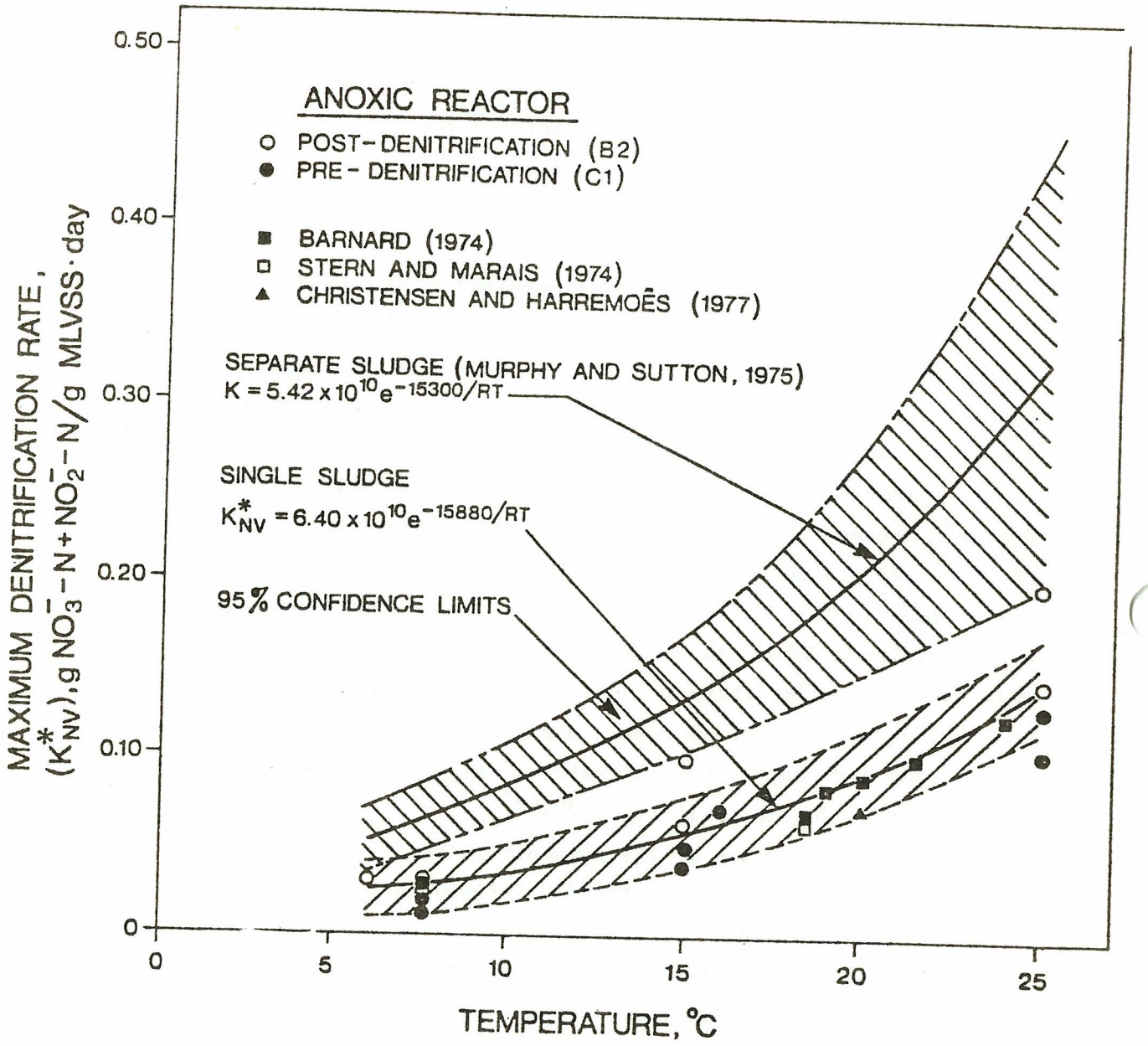


FIGURE 12. DENITRIFICATION RATES UNDER NON-CARBON LIMITING CONDITIONS

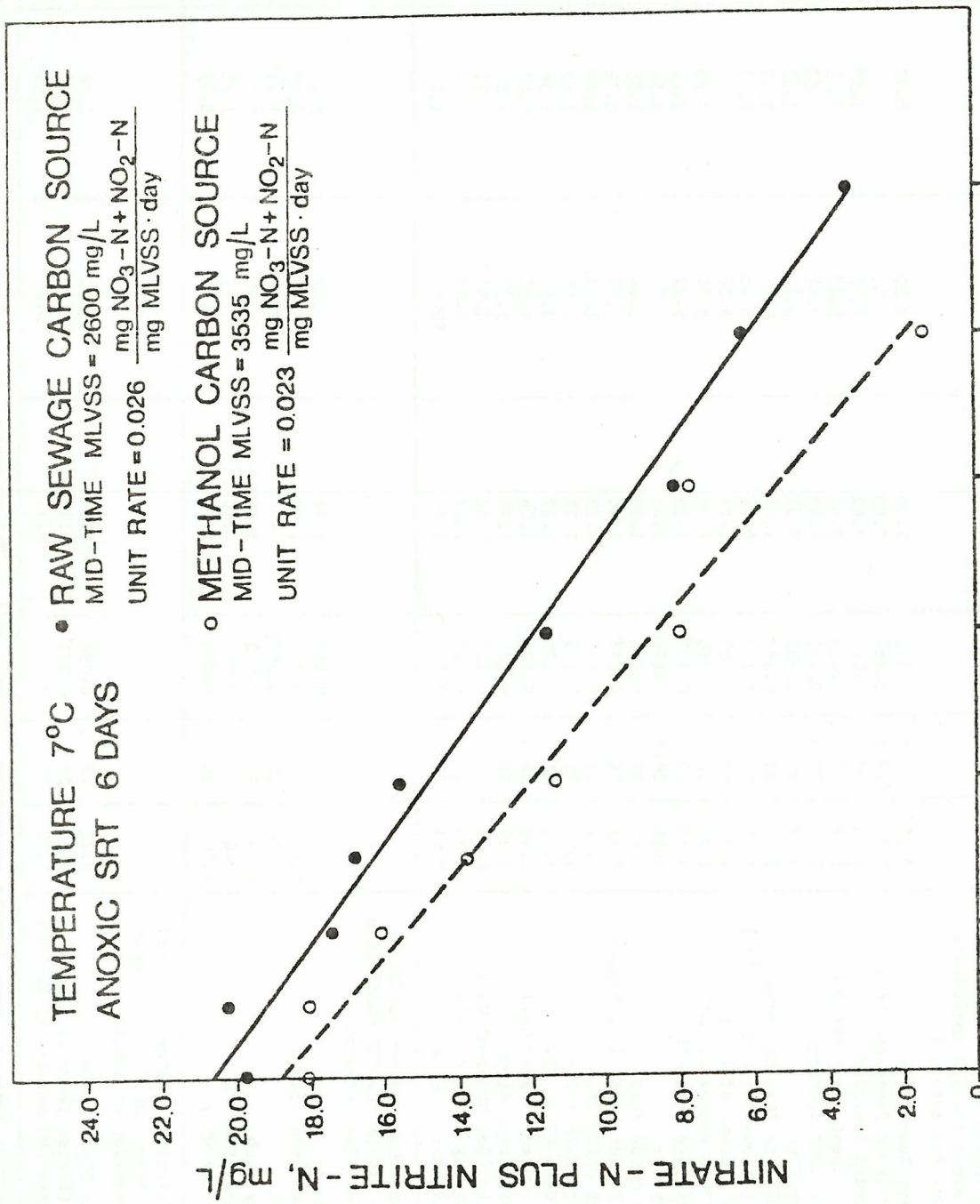


FIGURE 13. DENITRIFICATION RATES WITH METHANOL AND RAW SEWAGE AS CARBON SOURCES

TABLE I Denitrification Rates Using Industrial Wastes as the Carbon Source

Waste	Initial FOC:N	Temp. (°C)	DN Rate (d^{-1})	DN Rate Relative to MeOH Control Run on Same Day	Substrate Consumption Ratios		Comments*
					kg FCOD Consumed / kg NO_T-N Removed	kg FOC Consumed / kg NO_T-N Removed	
GROUP 1: The following wastes exhibited rates above the 95% confidence interval for methanol:							
Rieder Distillery Fusel Oils	3.19	20.5	0.331	2.38	2.22	0.77	
Pea Blanchwater (Food Processor 'A')	3.27	18.5	0.261	2.08	5.71		
Jordan Wines Sludge Centrate	2.70	20.5	0.207	1.62	7.30	2.28	1
Labatt's Brewery Spent Grain Extract	3.18	20	0.197	2.40	5.48	2.46	1
Molson's Brewery Last Runnings	2.53	20.5	0.191	1.49	6.67	1.83	
Molson's Brewery Wort	4.29	21	0.187	2.27	6.17	1.35	1
McGuinness Distillers Thin Stillage	2.71	21	0.184	1.44	6.07	2.18	1
Methanol Still Bottoms (Org. Manuf. 'A')	1.49	20	0.170	0.86	3.66	0.71	1
National Starch Process Effluent	2.97	18	0.160	1.54	3.26		
Tomato Sludge (Food Processor 'A')	1.72	18	0.160	1.31	2.54	0.80	
McGuinness Distillers Fusel Oils	3.17	20	0.159	1.29	5.32	1.46	1
Molson's Brewery Beer	4.16	20.5	0.159	1.41	8.57	2.54	1
Du Pont Organic Acids Waste	2.61	21	0.142	1.29	5.14	1.65	1
Spent Sulphite Liquor (Can. Int. Paper)	1.77	19	0.137	1.24	3.94	0.79	2,3
Domtar Packaging Whitewater	3.72	21	0.137	2.13	5.74	1.48	1,3
Vulcan-Cincinnati Methyl Fuel	4.07	21	0.135	1.22	6.18	1.83	
Celanese Light Ends (Tray 25)	3.48	21	0.129	1.17	5.23	1.36	
Methanol Heads (Ontario Paper Co.)	1.53	18	0.128	1.06	2.45	0.82	
Rieder Distillery Grape Slops	3.21	20	0.125	1.94	5.00	1.42	1
Acetic Acid Waste (Dow Chemical Co.)	1.76	20	0.123	0.62	3.87	1.71	1
Du Pont High Boiling Organic Waste	2.53	19	0.119	1.07	6.02	1.36	2
McGuinness Distillers Light Distillate	9.91	20	0.117	1.61	10.16	1.52	
GROUP 2: The following wastes exhibited rates within the 95% confidence interval for methanol:							
Jordan Wines Pomace Extract	3.43	19	0.112	1.74	5.69	2.6	1
Millhaven Fibres Glycol Waste	2.94	20	0.103	1.60	5.98	0.92	(DN is
METHANOL CONTROL	2.87	20	0.097†	-	5.41†	1.17†	-(Mean of
Molson's Brewery Trub	4.73	20	0.093	1.28	6.40	3.7	(14 Runs
Isopropanol Waste (Norwich)	4.40	20	0.090	1.40	3.64	1.82	2
Cos and Gris Cheese Whey	2.50	20	0.084	1.31	9.65	0.91	1
GROUP 3: The following wastes exhibited rates below the 95% confidence interval for methanol:							
Domtar Packaging Black Liquor	2.24	18	0.080	1.25	6.02	1.76	1,3
Waste Dextrose (Baxter Travenol Labs)	2.65	20	0.071	0.57	8.19	2.57	1
Formaldehyde Waste (University of Guelph)	3.62	20	0.042	0.37	6.21	1.38	

† Mean Value. *LEGEND: 1)Wastes cause nitrite production. 2)Waste adds to TKN concentration. 3)Waste adds colour to clarified effluent.

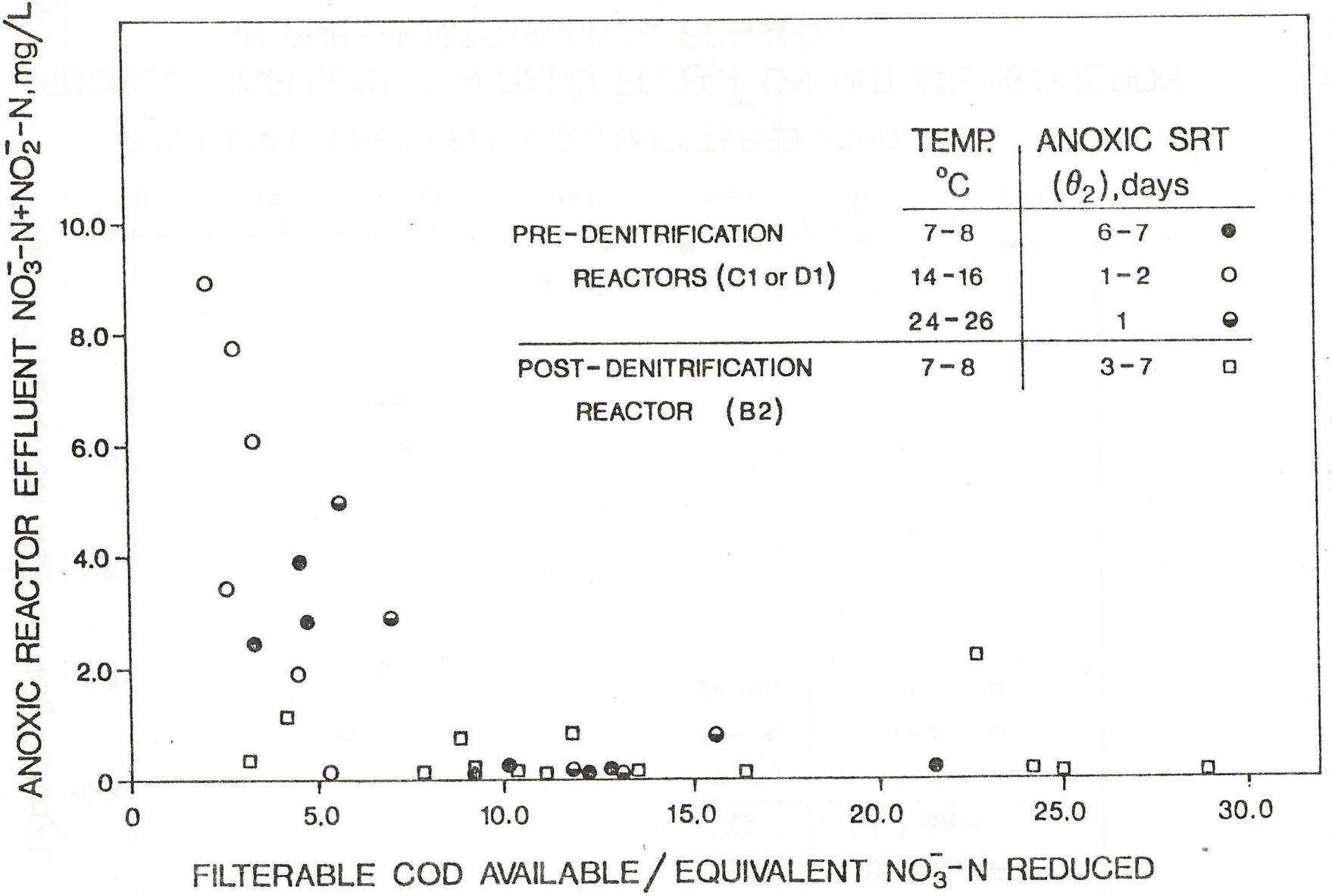


FIGURE 14. ORGANIC CARBON REQUIREMENTS IN POST AND PRE-DENITRIFICATION REACTORS

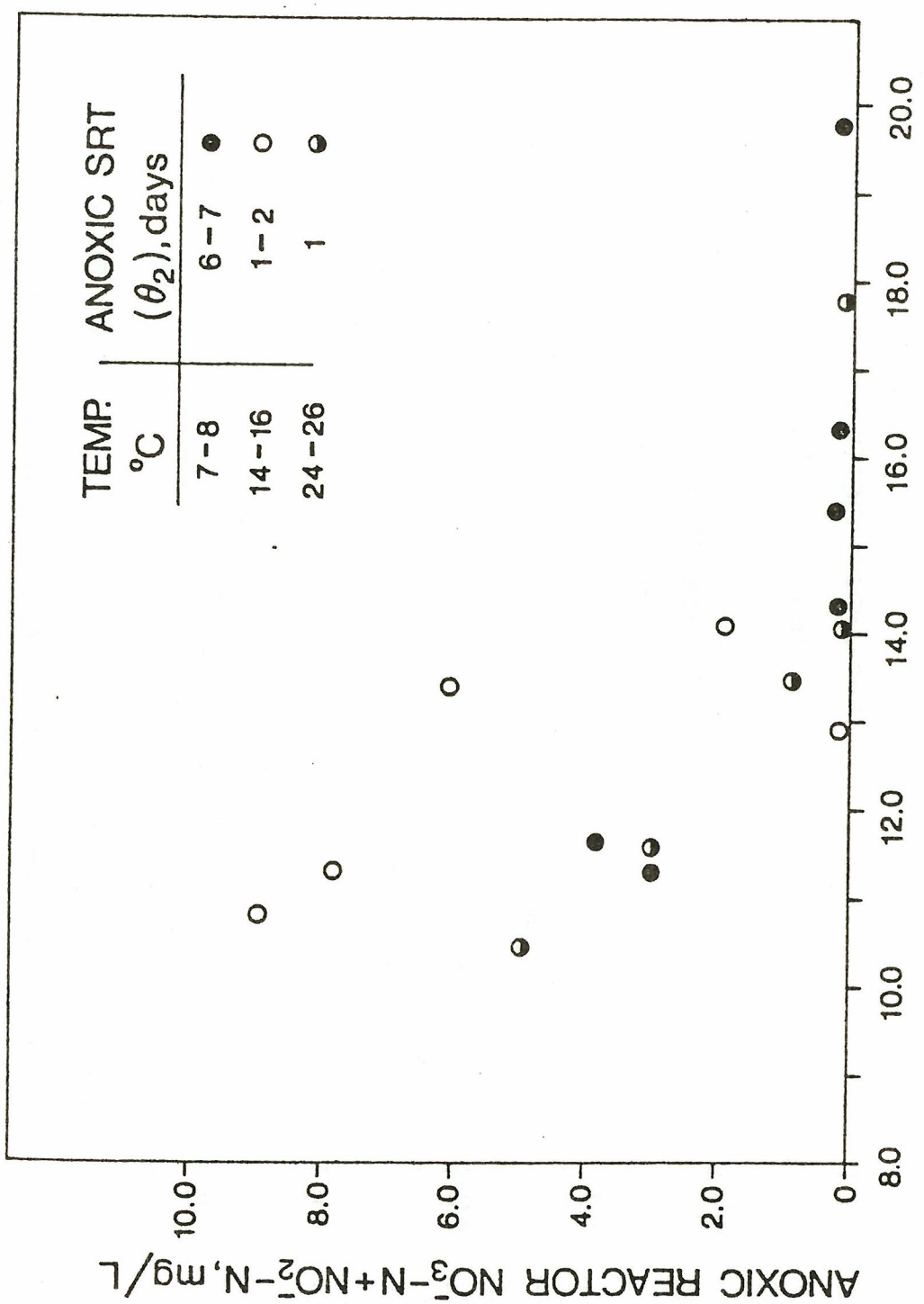


FIGURE 15. INFLUENT C:N RATIO EFFECT ON NITRATE REDUCTION IN PRE-DENITRIFICATION REACTOR

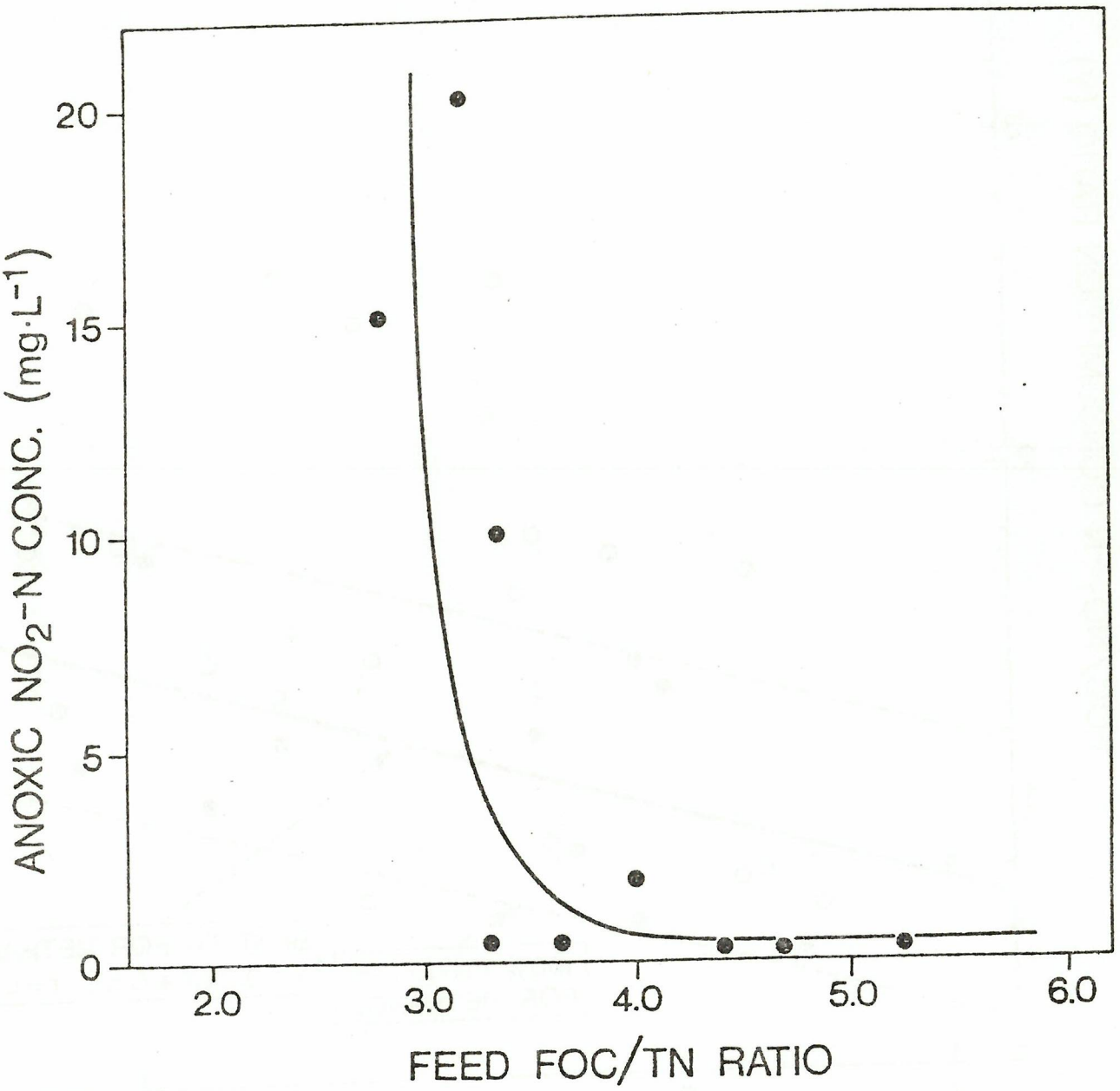


FIGURE 16. EFFECT OF FEED FOC/TN RATIO ON DENITRIFICATION

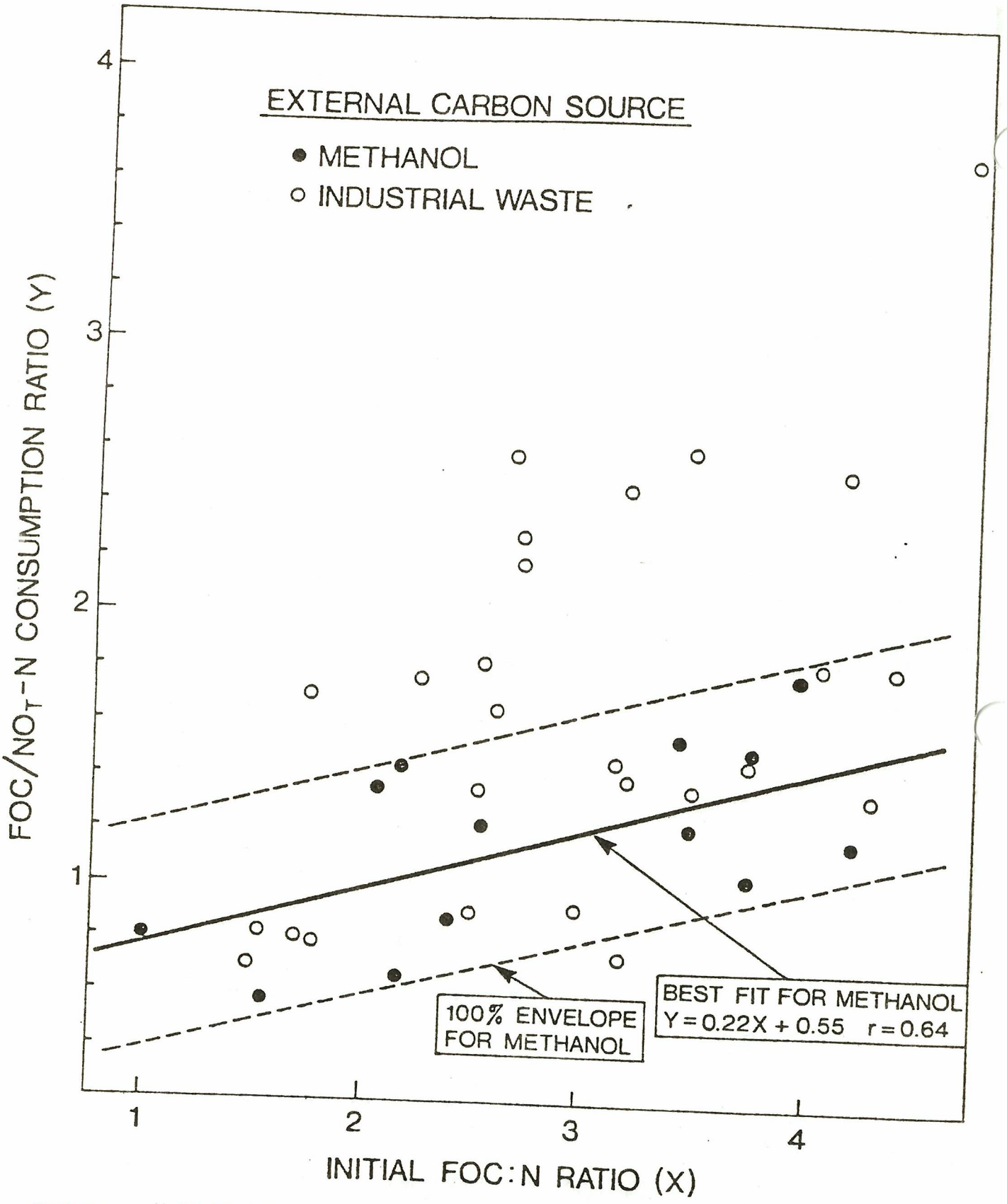


FIGURE 17. SUBSTRATE CONSUMPTION RATIO vs INITIAL FOC:N RATIO

REFERENCES

1. Shannon, E.E., and R.J. Rush, "Phosphorus Removal Treatability Studies at CFB Borden, Petawawa, Trenton and Uplands", Environment Canada, Environmental Protection Service Report No. EPS 4-WP-73-5, 1973.

2. Shannon, E.E., and R.J. Rush, "Development of Prediction Models for Chemical Phosphorus Removal - Vol. I", Canada-Ontario Agreement Research Report No. 68, 1977.

3. Stepko, W., "Phosphorus Removal Demonstration Studies at CFB Trenton, Phase II", Environment Canada, Environmental Protection Service Report No. EPS 4-WP-76-4, 1976.

4. U.S. EPA, Nitrogen Control Manual, Washington, D.C.

5. Bridle, T.R., D.C. Climenhage, and A. Stelzig, "Operating of a Full Scale Nitrification-Denitrification Industrial Waste Treatment Plant", J. Wat. Poll. Con. Fed., 51(1):127, 1979.

6. Wilson, R.S., K.L. Murphy, P.M. Sutton and S.L. Lackey, "Design and Cost Comparison of Biological Nitrogen Removal Systems", Presented at 51st Annual Wat. Poll. Con. Fed. Conference, Anaheim, California, 1978.

7. Haug, R.T., and P.L. McCarty, "Nitrification with Submerged Filters", J. Wat. Poll. Con. Fed., 44(11):2086, 1972.

8. Lawrence, A.W., and P.M. McCarty, "United Basis for Biological treatment Design and Operation", J. Sanitary Eng. Div., ASCE, 96(SA3):757, 1970.

9. Wuhrmann, K., "Effects of Oxygen Tension on Biochemical Reactions in Sewage Purification Plants", Advances in Biological Waste Treatment, Proceedings 3rd Conf. Biol. Waste Treatment, edited by W.W. Eckenfelder and J. McCabe, Pergamon Press, New York, p. 27, 1963.

10. Painter, H.A., "Microbial Transformation of Inorganic Nitrogen", Progress in Water Technology, 8, 415, 3, 1977.

11. Wilson, R.W., "Continuous Fixed Film Biological Nitrification and Denitrification of Wastewater", MSc Thesis, McMaster University, Hamilton, Ontario, 1975.

12. Sutton, P.M., "Suspended Growth Carbon Removal-Nitrification Systems", PhD Thesis, McMaster University, Hamilton, Ontario, 1976.

13. Buswell, A.M., T. Shiota, N. Lawrence, and I.V. Meter, "Laboratory Studies on the Kinetics of the Growth of Nitrosomonas with Relation to the Nitrification Phase of the BOD Test", Appl. Micro., 2, 21, 1954.

14. Deppe, K., and H. Engle, "Untersuchunger über Temperaturabhängigkeit der Nitratbildung durch Nitrobacter wingogradski Buch. bei ungehemmten under gehemmten Wachstum", Zentbl. Bakt. Parasitkde, 11, 113, 561, 1960.

15. Laudelout, H., and L. van Tichelen, "Kinetics of the Nitrite Oxidation by Nitrobacter winogradski", J. Bact., 79, 39, 1960.

16. Murphy, K.L., P.M. Sutton, R.W. Wilson, and B.E. Jank, "Nitrogen Control: Design Considerations for Supported Growth Systems", J. Wat. Poll Con. Fed., 49(4):549, 1977.

17. Huang, C.S., and N.E. Hopson, "Nitrification Rate in Biological Processes", J. Envir. Eng. Div., Proceedings of the Americal Soc. of Civil Engineers, 100, 409, 1974.

18. Downing, A.L., T.G. Tomlinson, and G.A. Truesdale, "Effect of Inhibitors on Nitrification in the Activated Sludge Process", Institute of Sewage Purification Journal and Proceedings, 537, 1964.

19. Hockenbury, M.R., and C.P.L. Grady, "Inhibition of Nitrification - Effects of Selected Organic Compounds", J. Wat. Poll. Con. Fed., 49(5):768, 1977.

20. Anthonisen, A.C., R.C. Loehr, T.B.S. Prakasam, and E.G. Srinath, "Inhibition of Nitrification by Ammonia and Nitrous Acid", J. Wat. Con. Fed., 48(5):835, 1976.

21. Kiff, R.J., "The Ecology of Nitrification/Denitrification in Activated Sludge", J. Wat. Poll. Con. Fed., 71, 475, 1975.

22. Heidman, J.A., "An Experimental Evaluation of Oxygen and Air Activated Sludge Nitrification Systems With and Without pH Control", U.S. EPA Report for Contract No. 68-03-0349, 1975.

23. Downing, A.L., H.A. Painter, and G. Knowles, "Nitrification in the Activated Sludge Process", J. Inst. Sew. Purif., 130, 1964.

24. Sutton, P.M., T.R. Bridle, W.K. Bedford, and J. Arnold, "Nitrification and Denitrification of an Industrial Wastewater: Demonstration of Process Capability", Presented at Biological Nitrification/Denitrification of Industrial Wastes Workshop, Wastewater Technology Centre, Burlington, Ontario, October, 1977.

25. Parker, D.S., Brown and Caldwell, Personal Communication to P.M. Sutton, 1977.

26. Wong-Chong, G.M., and R.C. Loehr, "Kinetics of Microbial Nitrification as Applied to the Treatment of Animal Waste - Ammonium-Nitrogen Oxidation", AIChE Symposium Series No. 145, 71, 1975.

27. Sutton, P.M., K.L. Murphy, and B.E. Jank, "Nitrification Systems with Integrated Phosphorus Precipitation", Presented at the Pollution Control Association of Ontario, Toronto, Ontario, April, 1977.

28. Poduska, R.A., and J.F. Andrews, "Dynamics of Nitrification in the Activated Sludge Process", J. Wat. Poll. Con. Fed., 47, 1599, 1975.

29. Sutton, P.M., K.L. Murphy, and B.E. Jank, "Kinetic Studies of Single Sludge Nitrogen Removal Systems", IAWPR Conference on Advanced Treatment and Reclamation of Wastewater, Johannesburg, Republic of South Africa, 1977.

30. Adams, C.E. Jr., and W.W. Eckenfelder, "Nitrification Design Approach for High Strength Ammonia Wastewaters", J. Wat. Poll. Con. Fed., 49(3):413, 1977.

31. Sutton, P.M., and D.S. Parker, Discussion of "Nitrification Design Approach for High Strength Ammonia Wastewaters" by C.E. Adams, Jr., and W.W. Eckenfelder, J. Wat. Poll. Con. Fed., in press, 1977.

32. Delwiche, C.C., "Denitrification", Symposium on Inorganic Nitrogen Metabolism, edited by McElroy & Glass, John Hopkins Press, Baltimore, Maryland, p. 233, 1956.

33. Myers, J., and F.A. Matsen, "Kinetic Characteristics of Warburg Manametry", Arch. Biochem. & Biophys., 55, 373, 1955.

34. Schmidt, B., and W.D. Kampf, "Über den Einfluss des Sauerstoffs auf die Denitrifikationsleistung von Pseudomonas fluorescens", Arch. Hyd. Bakt., 146, 171, 1962.

35. McCarty, P.L., L. Beck, and P. St. Amant, "Biological Denitrification of Wastewaters by Addition of Organic Materials", Proceedings of 24th Industrial Waste Conference, 1271, Purdue University, Indiana, 1969.

36. Monteith, H.M., T.R. Bridle, and P.M. Sutton, "Industrial Waste Carbon Sources for Biological Denitrificaton", Canada-Ontario Agreement Research Report, in press, 1977.

37. McCarty, P.L., "Nitrification-Denitrification by Biological Treatment", Correspondence Conference on Denitrification of Municipal Wastes, University of Massachusetts, Mass., 1973.

38. Chang, J.P., and J.G. Morris, "Studies on the Utilization of Nitrate by Micrococcus denitrificans", J. gen. Microbiol., 29, 301-310, 1962.

39. Dawson, R.N., and K.L. Murphy, "Factors Affecting Biological Denitrification of Wastewater", Advances in Water Pollution Research, edited by S.H. Jenkins, Pergamon Press, Oxford, England, 1973.

40. Mulbarger, M.C., "Nitrification and Denitrification in activated Sludge Systems", J. Wat. Poll. Con. Fed., 43(10):2040, 1971.

41. Christensen, M.H., and P. Harremoës, 2Biological Denitrification of Sewage: A Literature Review", Progress in water Technology, 8, 415, 509, 1977.

42. Barnard, J.L., "Cut P and N Without Chemicals", Water and Wastes Engineering, 11(7):33, 1974.

43. Sutton, P.M., K.L. Murphy, and B.E. Jank, "Design Considerations for Integrated Nutrient Removal Systems", Paper Prepared for consideration for IAWPR Conference, Stockholm, Sweden, June, 1978.

44. Sutton, P.M., K.L. Murphy, and R.N. Dawson, "Low Temperature Biological Denitrification of Wastewater", J. Wat. Poll. Con. Fed., 47(1), 122, 1975.

45. Murphy, K.L., and P.M. Sutton, "Pilot Scale Studies on Biological Denitrification", Progress in Water Technology, 7, 317, 1975.

46. Adams, C.E. Jr., and W.W. Eckenfelder, Process Design Techniques for Industrial Waste Treatment, Enviro Press, Nashville, Tennessee, p. 221, 1974.

47. Moore, S.F., and E.D. Schroeder, "An Investigation of the Effects of Residence Time on Anaerobic Bacterial Denitrification", Water Research, 4, 685, 1970.

48. Requa, D.A., and E.D. Schroeder, "Kinetics of Packed Bed Denitrification", J. Wat. Poll. Con. Fed., 45(8):1696, 1973.

49. Harremoës, P., "The Significance of Pore Diffusion to Filter Denitrification", J. Wat. Poll. Con. Fed., 48(2):377, 1976.

50. Bridle, T.R., D.C. Climenhage, and A. Stelzig, "Start-up of a Full Scale Nitrification-Denitrification Treatment Plant for Industrial Waste", 31st Purdue Industrial Waste Conference, Purdue Unviersity, Indiana, 1976.

51. Sutton, P.M., K.L. Murphy, B.E. Jank, and B.A. Monaghan, "Efficacy of Biological Nitrification", J. Wat. Poll. Con. Fed., 47(11):2665, 1975.

52. Lawrence, A.W., and C.G. Brown, "Design and Control of Nitrifying Activated Sludge Systems", J. Wat. Poll. Con. Fed., 48(7):1779, 1976.

53. Patterson, J.W., and R.A. Minear, "Pollution Control in the Explosives Industry", Proceedings 29th Industrial Waste Conference, Purdue Unitersity, Indiana, 1974.

54. Hutton, W.C., and S.A. La Rocca, "Biological Treatment of Concentrated Ammonia Wastewaters", J. Wat. Poll. Con. Fed., 47(5):989, 1975.

55. Murphy, K.L., P.M. Sutton, and B.E. Jank, "Dynamic Nature of Nitrifying Biological Suspended Growth Systems", Progress in Water Technology, 9(1):279, 1977.

56. Stensel, H.D., C.S. McDowell, and E.D. Ritter, "An Automated Biological Nitrification Toxicity Test", J. Wat. Poll. Con. Fed., 48(10):2343, 1976.

57. Wong-Chong,G.M., S.C. Caruso, and T.C. Patarlis, "An Evaluation of EPA Recommended Technology for the Treatment and Control of Wastewaters from By-Product Coke Plants - Alternate 2", Proceedings of 9th Mid-Atlantic Industrial Waste Conference, Bricknell University, Lewisburg, Pennsylvania, August, 1977.

58. Smith, A.G., "Nitrogen Removal: An Evaluation of Full Scale Biological Systems", "Canada-Ontario Agreement on Great Lakes Water Quality, Proceedings High Quality Effluent Seminar, Toronto, Ontario, December, 1975.

59. Dijkstra, F., and V.E.A. Baenens, "Biological Treatment of Wastewater of a Chemical Industry with Complete Nitrogen Removal", Proceedings Biological Nitrification/Denitrification of Industrial Wastewater Workshop, Wastewater Technology Centre, Burlington, Ontario, October, 1977.

60. Jeris, J.S., R.W. Owens, R. Hickey and F. Flood, "Biological Fluidized Bed Treatment for BOD and Nitrogen Removal", J. Wat. Poll. Con. Fed., 49(5):816, 1977.

61. Antonie, R.L., Fixed Biological Surfaces - Wastewater Treatment, CRC Press Inc., Celveland, Ohio, 1976.

62. English, J.N., C.W. Carry, A.M. Masse, J.B. Pitkin, and F.D. Dryden., "Denitrification in Granular Carbon and Sand Columns", J. Wat. Poll. Con. Fed., 46(1):28, 1974.

63. Sutton, P.M., B.E. Jank, and B.A. Monaghan, "Single Sludge Nitrogen Removal Systems", Canada-Ontario Agreement Research Report No. 88, 1979.

64. Bridle, T.R., W.K. Bedford, and B.E. Jank, "Biological Nitrogen Control of Coke Plant Wastewaers, in press, 1979.

65. Wilson, R.W., K.L. Murphy, and P.M. Sutton, "Design and Cost Comparison of Biological Nitrogen Removal Systems", Presented at 51st Annual Wat. Poll. Con. Fed. Conference, Anaheim, California, 1978.

FUNDAMENTALS OF PHYSICAL-CHEMICAL PROCESSES FOR THE REMOVAL OF NITROGEN COMPOUNDS FROM WASTEWATER

Part 1 - Process Fundamentals & Costs
Part 2 - Design Example

By
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FUNDAMENTALS OF PHYSICAL CHEMICAL PROCESSES FOR THE REMOVAL OF NITROGEN COMPOUNDS FROM WASTEWATER

INTRODUCTION

Various compounds of nitrogen which may occur in raw or treated wastewater are:

- Organic Nitrogen
- Ammonia Nitrogen--present either as the ammonium ion (NH_4^+) or as dissolved ammonia gas (NH_3)
- Nitrate Ion (NO_3^-)
- Nitrite Ion (NO_2^-)

With the notable exception of reverse osmosis technology most physical chemical methods of nitrogen removal are highly specific for nitrogen compounds in only a single form. Therefore, before selection and design of an appropriate removal process can be undertaken, the designer must be cognizant not only of the amount of nitrogen to be removed but its particular form as well. Such knowledge gained from actual testing of the wastewater to be treated is vital and it should be axiomatic that such information be obtained prior to design. For the purposes of today's discussion, however, concentration of nitrogen compounds typically found in domestic wastewater can serve as a basis for comparison of potential removal methods. Table 1 lists typical ranges of nitrogen concentrations found in domestic wastewaters as reported by five separate references (1, 2, 3, 4, 5).

TABLE 1

TYPICAL RANGES OF NITROGEN CONCENTRATIONS FOUND IN RAW DOMESTIC WASTEWATER

NITROGEN FORM	CONCENTRATION IN mg/l AS N REFERENCES				
	1	2	3	4	5
ORGANIC N	8 to 20	8 to 35	20	10 to 15	7 to 22
AMMONIA N	12 to 35	12 to 50	35	25 to 35	4 to 35
NITRATE N	0 to 0.5	0	NR	0	NR
NITRITE N	0 to 2.0	0	NR	0	NR
TOTAL N	20 to 55	20 to 85	55	35 to 55	11 to 57

REF. 1 — Public Works, May 1973
REF. 2 — Wastewater Engineering, Metcalf and Eddy, McGraw Hill Pub. Co. 1972
REF. 3 — MOP No. 8, WPCF, 1977
REF. 4 — Manual of Treatment Processes, Environmental Services Corp., 1968
REF. 5 — The Effect of Industrial Wastes on Sewage Treatment, New England Interstate WPCC, 1965

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Physical chemical processes are seldom if ever used for nitrogen removal from raw sewage; rather they generally follow preceding unit processes such as clarification and biological treatment. These precedent processes will in some cases remove nitrogen and will in most cases at least alter its form. Therefore, in addition to knowledge of raw wastewater characteristics, the designer of physical chemical nitrogen removal processes must be in a position to know what specific changes in both amounts and form of nitrogen will have occurred as a result of preceding processes.

Typically, well-operated activated sludge secondary processes will produce effluent having an organic nitrogen concentration of about 2 milligrams per litre (mg/l). Trickling filter type secondary processes appear to be less efficient in organic nitrogen removal with effluent ranges from 4 to 7 mg/l being recorded in Dallas, Texas and Orange County Sanitation Districts (6) both in the USA. Conversion from the ammonia form to the nitrate form of nitrogen often occurs in biological secondary treatment processes whether intended to or not.* Such conversion is largely a function of design and operational criteria and is dependent on several factors including air supply, temperature, and sludge age. Attempting to design from "typical" ammoniacal nitrogen changes through secondary treatment is, therefore, neither prudent nor accurate. Sufficient to say that the designer of physical chemical processes for ammonia removal, which are to be located downstream of biological secondary treatment, must be cognizant of the design and operation of preceding processes. These in conjunction with the standard which is to be met will play a major roll in the selection of appropriate ammonia removal processes.

In addition to its ubiquitous nature, ammonia in wastewater is a "changeling." Not only is it relatively easily oxidized from the ammonia form to nitrate, but it may also take the form of a dissolved gas, an ammonium ion in true solution, or both depending on pH and temperature. At a pH of 7 or less, only ammonium ions in true solution are present. At a pH of 12, only dissolved ammonia gas is present. At pH values between 7 and 12, both forms may be present. Since specific ammonia removal methods will likely be dependent upon the form of the ammonia present, pH and

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* Perversely, nitrification seems to occur if not intended and not to occur if intended!

temperature become additional important factors to consider. The effects of pH and temperature on the distribution of ammonia and ammonium ion in water are shown in Figure 1.

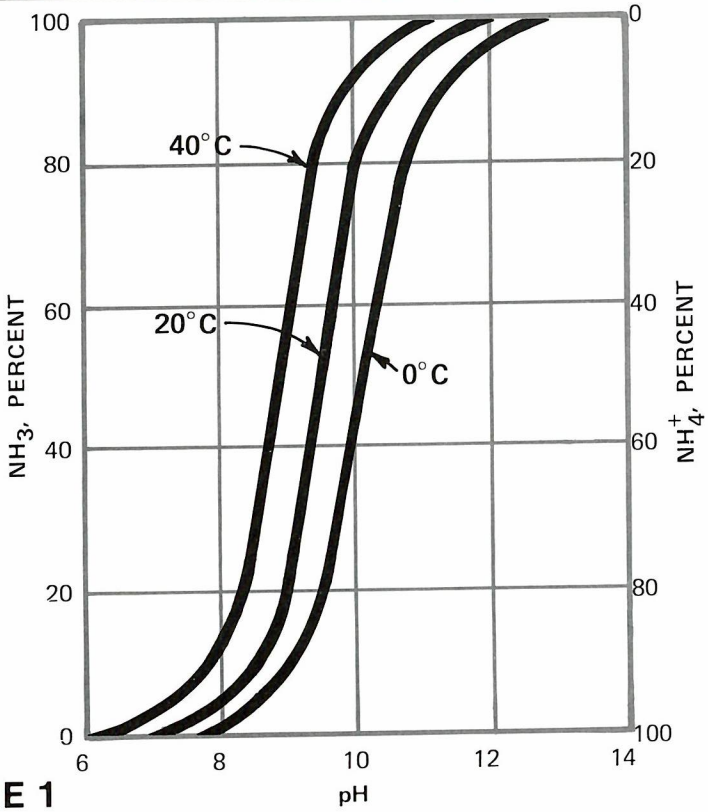
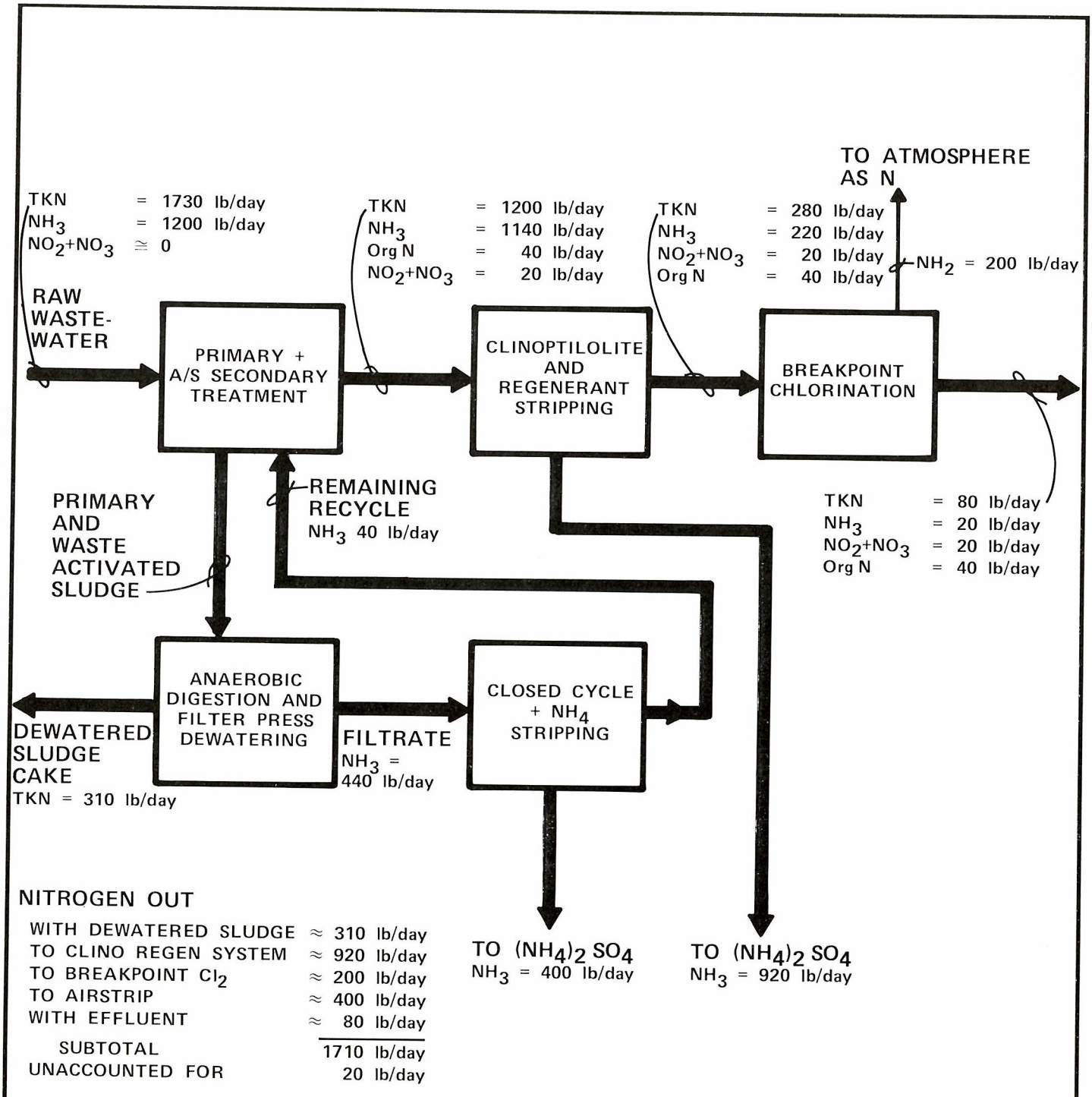


FIGURE 1

EFFECTS OF pH AND TEMPERATURE ON DISTRIBUTION OF AMMONIA AND AMMONIUM ION IN WATER

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In addition to the amount and form of nitrogen present in raw wastewaters, and the type and operational modes of processes which precede nitrogen removal, an additional important facet for consideration is impact of recycle streams on downstream nitrogen removal processes. This important factor is unfortunately often overlooked by designers and, if overlooked, can have a very significant deleterious effect. The potential magnitude of the impact of recycles is illustrated by Figure 2, which represents a mass nitrogen balance for the 4.83 million gallon (US) per day (mgd) Tahoe-Truckee Sanitary Authority Plant near Lake



NITROGEN OUT

WITH DEWATERED SLUDGE	≈	310 lb/day
TO CLINO REGEN SYSTEM	≈	920 lb/day
TO BREAKPOINT Cl_2	≈	200 lb/day
TO AIRSTRIP	≈	400 lb/day
WITH EFFLUENT	≈	80 lb/day
SUBTOTAL		1710 lb/day
UNACCOUNTED FOR		20 lb/day

FIGURE 2

SIMPLIFIED NITROGEN MASS BALANCE FOR TAHOE-TRUCKEE SANITARY AUTHORITY AWT PLANT AT 4.83 MGD (U.S.) (18,286 M^3/day) FLOW RATE

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Tahoe in California, USA. Figure 2 is based on actual nitrogen data collected over the past 2 years of operation of this facility. Note that the plant flow scheme includes high rate anaerobic digestion followed by sludge dewatering accomplished by filter press. On the average, filtrate resulting from filter pressing of digested sludge is shown to contain about 33 percent as much ammonia nitrogen as the raw incoming wastewater; therefore, neglecting this sizable sidestream could have resulted in undersizing of ammonia removal processes by a full 25 percent.

REMOVAL OF NITROGEN

The methods for removal of nitrogen from wastewater presented in this paper are strictly limited to those physical and chemical in nature. The principal processes that have been considered or used for full scale operations include:

- Air Stripping
- Breakpoint Chlorination
- Ion Exchange

Air Stripping

The first nitrogen removal process used on a plant scale in wastewater treatment was the air stripping process employed at the South Lake Tahoe plant.

The principles of the process are simple. Ammonia nitrogen present in secondary effluent can be converted from ammonium ion (NH_4^+) to ammonia gas (NH_3) by raising the wastewater pH. The gaseous ammonia can then be removed by passing the high pH wastewater through a stripping tower, where the formation of water droplets in the presence of a large air flow strips the ammonia gas from solution. This process is compatible with lime treatment for phosphorus removal in that the lime treated and clarified wastewater is at or near the optimum pH condition for air stripping.

The stripping process consists of raising the pH of the wastewater to 10.8 to 11.5 (often done with lime to simultaneously remove phosphorus); formation and reformation of water droplets in a stripping tower; and providing air-water contact and droplet agitation by circulation of large quantities of air through the tower. The towers used for ammonia stripping may closely resemble conventional cooling towers.

Process Advantages. One of the most significant advantages of this process is its simplicity. Wastewater having a high pH is pumped to the top of a stripping tower. A fan located

on top of the tower draws air through the packing material, and ammonia is stripped from the water droplets. The primary control necessary is maintenance of the proper influent wastewater pH.

Another advantage of the ammonia stripping process is that it does not create any residual solids requiring disposal and does not have any significant recycle flow streams that impact other processes. The ammonia removed is discharged to the atmosphere.

There has been concern, however, over the potential for creation of an air pollution problem by the ammonia thus stripped. Scavenging of the ammonia from the atmosphere could, in turn, find its way back to receiving waters.

The concentration of ammonia in the stripping tower discharge is only about 6 milligrams per cubic metre (mg/m^3) for domestic wastewaters (at an air flow of 550 cubic feet per gallon and at an ammonia concentration of 23 mg/l in the tower influent). As the odor threshold of ammonia is 35 mg/m^3, the process does not present a pollution problem in this respect. The ammonia discharged to the atmosphere is a stable material which is not oxidized to nitrogen oxides in the atmosphere. The natural production and release of ammonia as part of the earth's natural nitrogen cycle is about 50,000,000,000 tons per year. Roughly 99.9 percent of the atmosphere's ammonia concentration is produced by natural biological processes (7). There is a large turnover of ammonia in the atmosphere with the total ammonia content being displaced once per week on the average. Ammonia is returned to the earth through gaseous deposition (60 percent); aerosol deposition (22 percent); and precipitation (18 percent). Ammonia is not considered an air pollutant because there are no known public health implications and because it is a natural consituent of the atmosphere derived almost entirely from natural sources. Therefore, the process cannot be dismissed from consideration because of air pollution.

Ammonia may be washed out from air by rainfall but not by snowfall. The natural background concentration of ammonia in the atmosphere varies from .001 to .02 mg/m^3. In rainfall, the natural background ranges 0.01 mg/l to 1.0 mg/l (8). The amount of ammonia in rainfall is directly related to the concentration of ammonia in the atmosphere. Thus, an increase in the ammonia in rainfall would occur only in that area where the stripping tower discharge increases the natural background ammonia concentration in the atmosphere.

Calculations for the ammonia washout which would occur with a rainfall rate of 3 mm/hr (0.12 in/hr) have been made for the Orange County, California project with the following results: (9)

Distance From Tower ft	Distance From Tower m	Peak Rainfall Ammonia Concentration, mg/l
300	91	60
1,000	305	18
1,600	488	11
3,200	975	5
16,000	4,877	0.5

The concentrations of ammonia in the rainfall would approach natural background levels within 16,000 ft (4.8 km) of the tower. The ultimate fate of the ammonia which is washed out by rainfall within this 16,000 ft (4.8 km) downwind distance depends on the nature of the surface upon which it falls. Most soils will retain the ammonia. That portion which lands on paved areas or directly on a stream will appear in the runoff from that area. Unless the stripping tower is located upwind in close proximity to a lake or reservoir, the direct return of ammonia to the aquatic environment by atmospheric washout should not make a significant contribution to the total ammonia discharged to the aquatic environment.

Process Disadvantages. The process also has its limitations and operating problems. The major limitation is the effect of temperature on process efficiency. As the air temperature drops, the efficiency of the process also drops. For example, at South Tahoe, the stripping tower removed about 90 percent of the ammonia in warm weather (20 degrees C air temperature), but only about 75 percent of the ammonia at a temperature of 10 degrees C. The tower was inoperable (due to freezing problems) when the air temperature fell very far below freezing. It is not practical to heat the large volumes of air required for the stripping process, as used at South Tahoe. Heating of the liquid is also not feasible because any heat contained in the tower influent would be dissipated in the top few inches of its downward passage through the tower.

A second problem is that of calcium carbonate scale deposition on the tower fill and structural members. The scale which forms can be removed easily and economically if the tower is designed for convenient removal of this scale. An example of this is the 15-mgd ammonia stripping process at the Orange County, California, "Water Factory 21." There, the tower packing can be easily removed for

cleaning, although no significant scaling problems have resulted to date (10). Scale was also found to be an insignificant problem at the Windhoek, South Africa, Plant, where a soft, easily removed scale was encountered (11). At the Blue Plains Pilot Plant, on the other hand, a hard scale which was extremely difficult to remove was encountered (12). The hardness of the scale at Blue Plains was affected by operating pH, with a harder scale forming at pH 11.5 than at pH 10.8. the type of tower used (countercurrent or crossflow) also may influence the severity of the scaling problem.

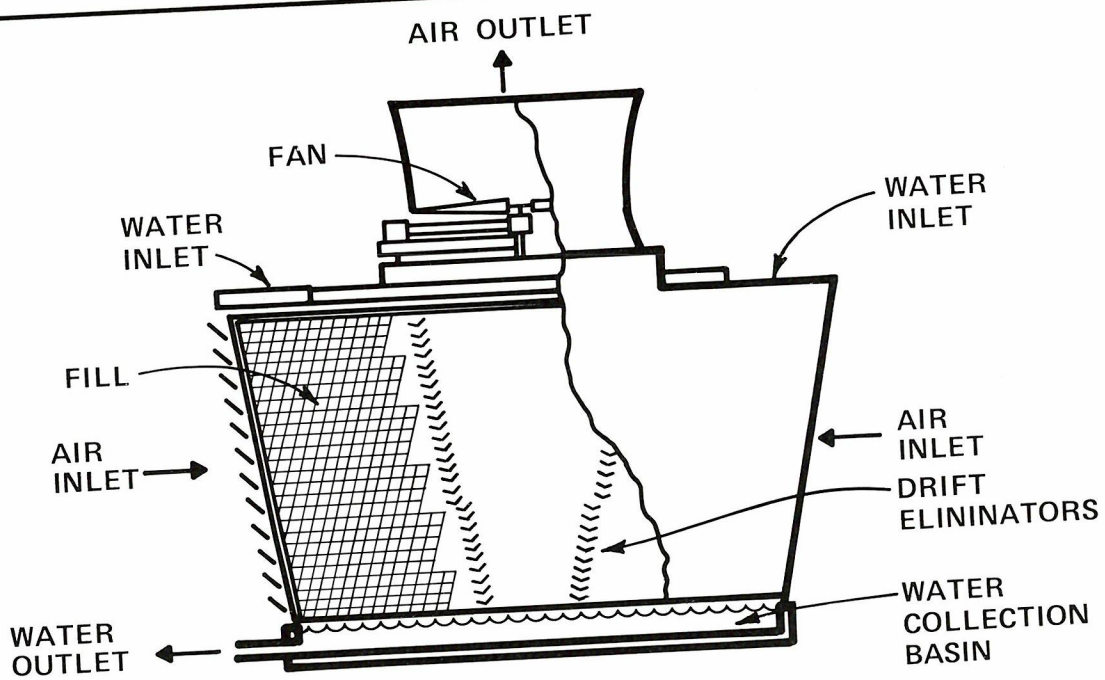
System Design Considerations. The principal factors affecting design and performance of ammonia stripping towers include the following:

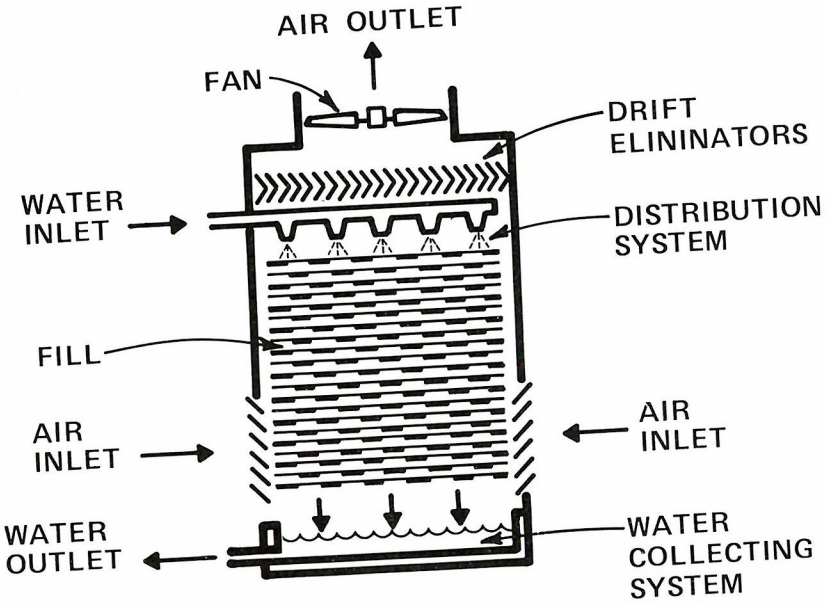
- Tower configuration
- pH
- Temperature
- Hydraulic loading
- Tower packing depth and spacing
- Air flow
- Scale control

Tower Configuration. There are two basic types of stripping towers: crossflow and countercurrent. Typical crossflow and countercurrent towers are shown in Figure 3.

With countercurrent towers, the entire airflow enters at the bottom and exits at the top while the water enters the top of the tower and drops through the packing material to the bottom. This type of tower has been found to be the most efficient. Crossflow towers also have the water entering the top of the tower, but the air is pulled into the tower through its sides throughout the height of the packing. The crossflow tower has been found to be more prone to scaling.

pH. The pH of the water entering the stripping tower has a major effect on the efficiency of ammonia removal. The pH must be raised sufficient to convert all of the ammonium ion to ammonia gas (see Figure 1). If phosphorous removal is required, the use of lime as the coagulant will generally enable the necessary pH elevation to be achieved concurrent with phosphorus removal. If pH elevation does not occur in some upstream processes, then the economics of the stripping process are adversely affected since the costs of pH elevation must then be incurred solely for ammonia stripping.





■ **FIGURE 3**

TYPES OF STRIPPING TOWERS

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Temperature. Air temperature is critical to process performance. The water temperature reaches equilibrium at a value near the air temperature in the top few inches of the stripping tower. As the water temperature decreases, the solubility of ammonia in water increases and it becomes more difficult to remove the ammonia by stripping. The amount of air per gallon must be increased to maintain a given degree of removal as temperature decreases. However, it is not practical to supply enough air to fully offset major temperature decreases. Data collected in pilot tests by EPA at the Blue Plains plant in Washington, D.C. illustrate the temperature effects and are shown in Figure 4 (9).

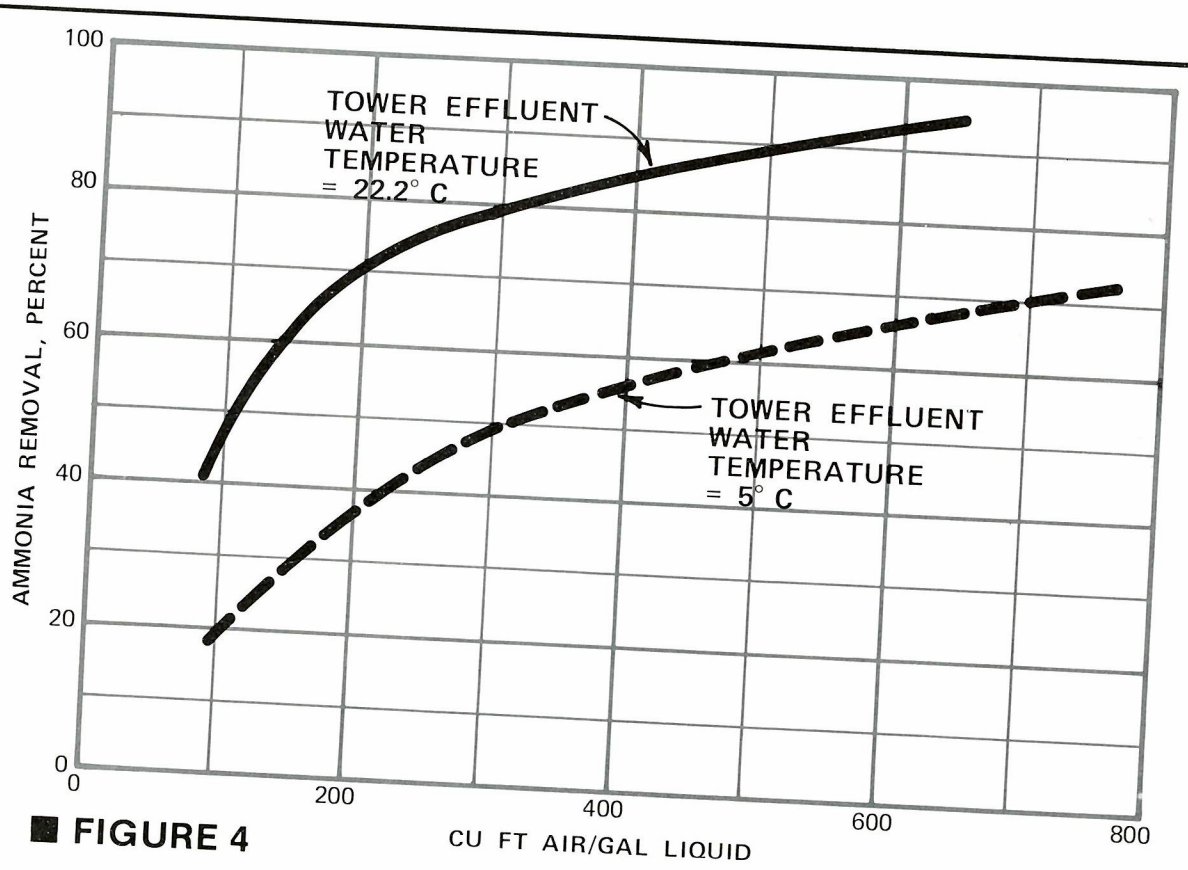


FIGURE 4

EFFECT OF TEMPERATURE ON AMMONIA REMOVAL EFFICIENCY OBSERVED BY BLUE PLAINS PILOT PLANT (12)

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Hydraulic Loading. The hydraulic loading rate of the tower is an important factor. This is typically expressed in terms of gallons/minute applied to each square foot of the plan area of the tower packing. When the hydraulic loading rate becomes too high, good droplet formation needed for efficient stripping is disrupted and the water begins to flow in sheets. If the rate is too low, the packing may not be properly wetted resulting in poor performance and scale accumulation. Operating results from South Tahoe and Orange County indicate that hydraulic loading rates should be in the range of 1.0 to 2.0 gpm/ft^2.

Tower Packing Depth and Material. The depth of tower packing required for maximum ammonia removal will depend on the tower packing selected. Most stripping tower designs are based on the use of an open, cooling tower-type packing (horizontal packing members spaced about 2 inches apart both horizontally and vertically) to minimize the power required to move adequate air quantities through the tower. If maximum ammonia removals are desired, tower packing depth should be at least 20 to 24 feet with this type of packing. Both wood (Lake Tahoe) and plastic packings (Orange County) have been used in full-scale towers. The smooth plastic surfaces appear to be one factor accounting for reduced calcium carbonate scaling at the Orange County facility. Plastic packing has an advantage in that it does not suffer from the delignification that occurs with wood at elevated pH values.

Air Flow. For a given height of tower packing, mass transfer relationships indicate that increased ammonia removal can be obtained by increased air flow. Because pressure drop increases with air flow rate, and consequently increases the power use, a practical upper limit is around 500 cfm or air per gpm of water (or an air pressure drop of 0.5 to 1.25 inches of water) (13). In moderate to cold climates, air stripping must be supplemented by a process such as breakpoint chlorination because the stripping efficiency becomes exceedingly low as the temperature of the water approaches freezing.

Scale Control. A factor which may have an adverse effect on tower efficiency is scaling of the tower packing resulting from deposition of calcium carbonate from the unstable, high pH water flowing through the tower. Scaling potential can be minimized by maximizing the extent of completion of the calcium carbonate reaction in the lime treatment step. Using a high level of solids recycle in the chemical clarification step

will ensure a more complete reaction. Another approach might be to eliminate CO_2 from the air supply to the tower. Also, a scale inhibiting compound added to the tower influent may offer a solution.

A further precaution that might be warranted in some instances is to design the towers so that the packing can be removed for cleaning, like at Orange County.

Design Criteria Summary. Typical design criteria for ammonia stripping towers are:

- Hydraulic loading--1 to 2 gpm/ft^2
- Air/water ratio--300-500 cu ft/gallon
- Air pressure drop--0.5 to 1.25 inches water
- Fan tip speed--9,000-12,000 fpm
- Packing depth--20-25 ft
- Packing spacing--2 to 4 inches horizontal and vertical
- Packing type--wood or plastic

Breakpoint Chlorination

When chlorine is added to a wastewater containing ammonia nitrogen, ammonia reacts with the hypochlorous acid formed to produce chloramines. Further addition of chlorine to the breakpoint converts the chloramines to nitrogen gas. The principal reactions of chlorine and ammonia in water are:

1. $NH_4^+ + HOCl \rightarrow NH_2Cl$ (monochloramine) $+ H_2O + H^+$

2. $2NH_2Cl + HOCl \rightarrow N_2 + 3HCl + H_2O$

Undesirable side reactions that can occur at unfavorable pH are:

3. $NH_2Cl + HOCl \rightarrow NHCl_2$ (dichloramine) $+ H_2O$

4. $NH_2Cl + HOCl \rightarrow NCl_3$ (nitrogen trichloride) $+ H_2O$

5. $NH_4 + 4HOCl \rightarrow NO_3^-$ (nitrate) $+ 4HCl + H_2O + H^+$

Breakpoint chlorination tests on domestic wastewaters at the Blue Plains Plant at Washington, D.C. indicated that 95 percent to 99 percent of the ammonia is converted to nitrogen gas and that no significant amount of nitrous oxide is formed (14). The quantity of chlorine required to achieve breakpoint was found to decrease with an increasing degree of treatment prior to the breakpoint process. The quantity of chlorine required for breakpoint chlorination of raw wastewater was found to be 10 parts by weight of Cl_2 to one part of NH_3-N. This ratio decreases to 9:1 Cl_2: NH_3-N for secondary effluents and 8:1 Cl_2:NH_3-N for lime clarified

and filtered secondary effluent. The Blue Plains tests found that the chlorine dose was minimized at pH values between 6.0 and 7.0. The minimum NO_3 production occurred at pH 5.0 (1.5 percent of the NH_3-N) to pH 8.0 (2.5 percent of the NH_3-N). Temperature did not affect the product distribution or the required chlorine dose in the range of 5 degrees C to 40 degrees C.

The use of chlorine produces an equivalent weight of hydrochloric acid which may depress the pH of the wastewater unless the natural alkalinity is adequate or a base such as sodium hydroxide is added. If the pH is too low, highly odorous nitrogen trichloride (NCl_3) forms and is an intolerable end product. If a base is used to prevent pH depression, the mixing of the wastewater, chlorine, and base must be extremely violent to avoid local areas of low pH which would generate NCl_3. Tests at Blue Plains showed that eductors do not give adequate mixing and do result in localized low pH regions where objectionable quantities of NCl_3 formed. Intense mixing is required. The use of sodium hypochlorite rather than chlorine does not depress the pH as much and reduces the amount of neutralizing agent that must be added.

The use of chlorine gas may produce more acid that can be neutralized by the wastewater. According to the Blue Plains study, 14.3 mg/l of alkalinity (as $CaCO_3$) are required to neutralize the acid produced by the oxidation of 1 mg/l of NH_3-N to N_2. A wastewater containing 20 mg/l of NH_3-N requires an alkalinity of about 286 mg/l if chlorine gas is used. Sodium hydroxide may be used for pH control if the wastewater is deficient in alkalinity. At Blue Plains, the NaOH quantity added to neutralize the acidic effects of chlorine addition was 0.9 to 1.7 pounds of NaOH per pound of chlorine, and corresponds to a final pH of from 6 to 7.9. At a 9:1 Cl/N ratio, 1.3 pounds of NaOH were required per pound of chlorine to maintain a pH of 7.

Process Advantages. The principal advantages of the breakpoint chlorination process are low capital costs, assurance of disinfection (particularly viral inactivation) and the nearly complete conversion of ammonia to nitrogen gas which can be harmlessly dissipated to the atmosphere. Furthermore, the breakpoint process is comparatively reliable with the proper control of pH and chlorine dose.

Process Disadvantages. Some of the disadvantages of breakpoint chlorination are:

- o High operating costs
- o Increased TDS content of wastewater
- o Complexity of control
- o Production of undesirable byproducts

Operating Cost. The high operating costs of the process are primarily due to the cost of chlorine, energy needed for mixing, and neutralizing agent, such as sodium hydroxide.

TDS Increase. The TDS (total dissolved solids) increase of the wastewater may not be a severe handicap to use of breakpoint chlorination, but is a factor that must be considered. An indication of the TDS increase is shown by the following relationships:

	mg/l TDS increase per mg/l NH_4-N removed
Breakpoint with chlorine gas	6.2
Breakpoint with sodium hypochlorite	7.1
Breakpoint with chlorine gas- Neutralization of all acidity with lime	12.2
Breakpoint with chlorine gas- Neutralization of all acidity with sodium hydroxide	14.8

The TDS increase may be particularly undesirable, if sodium hydroxide is the neutralizing agent, because of the increased sodium concentration in the final effluent. Sodium is of particular concern in drinking water because of its adverse health effects to people suffering from cardiac problems.

Process Control. Both South African (15) and American researchers (16) have reported that if a continuously functioning breakpoint chlorination process is to be a consistent and reliable unit process, the system must be capable of responding rapidly to changes in ammonia nitrogen concentration, chlorine demand, pH, alkalinity and flow.

Failure of the chlorine dosage control system to respond adequately to changes in process conditions can result in a substantial loss in nitrogen removal capability as

well as potentially adding significant overdoses of chlorine to the process. Overdoses of chlorine to the system are a direct waste of this chemical and result in increased difficulty of dechlorination, causing the direct discharge of high concentrations of chlorine residuals to the receiving water. Furthermore, overdoses may contribute to greater concentration of undesirable byproducts such as nitrogen trichloride and chlorinated organics.

A process control system successfully used for breakpoint chlorination at the Blue Plains pilot plant in Washington, D.C. The control of chlorine dosage was accomplished by a combination feed forward and feedback control loop. The feed forward component utilizes ammonia concentration and flow signals together with a manually selected multiplier to establish an approximate chlorine dosage to achieve breakpoint. The manually preselected multiplier corresponds to the Cl_2:NH_3-N ratio required for breakpoint and varies from values of about 8 to 10.

The feedback control loop involves measurement of the free chlorine residual in the process effluent. The level of free chlorine residual is compared to a setpoint value (usually 2-4 mg/l) and a signal is generated which provides a "trimming" of the chlorine dosage to the system.

Under most circumstances, lime or sodium hydroxide (NaOH) is added to the breakpoint process to neutralize a portion of the acidity generated in the chlorine addition to maintain the optimum pH. The lime or NaOH requirements at a particular installation are related to wastewater alkalinity, individual treatment processes employed prior to breakpoint chlorination as well as pH or alkalinity restrictions which might be imposed upon the breakpoint effluent.

pH control in the breakpoint chlorination process may be effectively achieved using a combination feed forward and feedback system. The feed forward component of the system involves pacing the lime or sodium hydroxide addition directly to the chlorine application rate. This accomplishes neutralization of a preselected portion of the chlorine acidity. A feedback loop is

employed to "trim" the system to a designated pH level. Accurate pH control of the process promotes efficient utilization of chlorine and can reduce the concentration of undesirable end products of the reaction.

Effective control over breakpoint chlorination requires utilization of accurate and reliable sensory equipment. In this regard two components deserve special attention namely the ammonia nitrogen and free chlorine monitoring elements.

The ammonia nitrogen and free chlorine monitoring devices used during the Blue Plains pilot testing of breakpoint chlorination were of the automated wet chemical analyzer type using colorimetric techniques to measure the ammonia nitrogen and free chlorine concentrations in small individual samples of the process influent. A continuous duty free chlorine residual analyzer was also successfully used at Blue Plains to provide a feedback signal for breakpoint chlorination control.

Production of Byproducts. Even the application of the best control techniques and maintenance of optimum pH, the breakpoint reaction will produce some residual nitrate and nitrogen trichloride in the treated water. Increased concentrations, however small, of either form of nitrogen may be undesirable or unacceptable.

Of current concern is the potential formation of chlorinated hydrocarbons by use of high chlorine dosage and residual chlorine concentrations. The opportunity for such reactions to occur is certainly available in the breakpoint process. Preliminary investigations have indicated that volatile chlorinated organics (such as chloroform) are not produced in concentrations greater than the limits proposed by EPA for drinking water. But there is much concern over the use of this process because of the potential formation many chlorinated organics which are not routinely monitored in wastewater effluents.

One additional concern is the quantity of chlorine used onsite, if the breakpoint process is used as the sole means of ammonia removal in a conventional secondary effluent. This may lead to consideration of onsite generation of chlorine or sodium hypochlorite, which adds greatly to the complexity of the overall process.

System Design Considerations. The factors of most concern in design and performance of the breakpoint process are:

- o Mixing
- o Contact time
- o Chlorine dosage
- o pH
- o Basin configuration

Mixing. At the point of addition of chlorine, mixing must be intense to assure uniform pH and chlorine residual. A velocity gradient of 900 to 1,000 ft/sec/ft should be provided for this purpose. A short detention time is desirable for this purpose to minimize the mixer horsepower required.

Time. After initial mixing, a contact time of 30 minutes is desirable for completion of the reaction. This is best accomplished in a plug flow reactor with a serpentine flow pattern.

Cl_2 Dosage. Design of the chlorine feed facilities should be such that a weight ratio of chlorine:ammonnia nitrogen (Cl_2:NH_4-N) can be maintained at 10:1 under the maximum ammonia concentration foreseen.

pH. Facilities must be provided for feeding an acid neutralizing compound such as lime or sodium hydroxide to maintain the wastewater pH at 7.0.

Selective Ion Exchange

The selective ion exchange process derives its name from the use of zeolites which are selective for ammonia relative to calcium, magnesium, and sodium. The zeolite currently favored for this use is clinoptilolite, which occurs naturally in several extensive deposits in the western United States. The order of preference for the various cations by clinoptilolite is:

$$Cs^{+}>Rb^{+}>K^{+}>NH_4^{+}>Ba^{++}>Sr^{++}>Na^{+}>Ca^{++}>Fe^{+3}>Fe^{+3}>Al^{+3}>Mg>Li$$

The clinoptilolite is crushed and sieved to obtain a 20 x 50 mesh size. Ammonia is removed by passing the wastewater through a bed of clinoptilolite at a rate of about 6-10 bed volumes per hour. Ammonia removal of 96 percent may be expected with influent ammonia nitrogen concentrations of about 20 mg/l.

The ammonia capacity of the clinoptilolite is nearly constant over the pH range of four to eight, but diminishes rapidly outside this range. Wastewater composition has an

effect on the ammonia exchange capacity. For relatively constant influent ammonia concentrations, the ammonia exchange capacity decreases sharply with increasing concentrations of competing cations (such as calcium and magnesium). Ammonia removal to residual levels below 0.5 mg/l ammonia nitrogen is technically feasible, but requires shorter service cycles and greater regeneration requirements. Flow rates in the range of 7.5 to 15 bed volumes per hour indicate no effect on ammonia effluent values.

After about 150-200 bed volumes of normal strength municipal waste have passed through the bed, the capacity of the clinoptilolite has been used to the point that ammonia begins to leak through the bed. At this point, the clinoptilolite must be regenerated so that its capacity to remove ammonia is restored.

Regeneration. The key to the applicability of this process is the method of handling the spent regenerant. The exchange media is regenerated by passing concentrated salt solutions through the exchange bed when the ammonia concentration has reached the maximum desirable level. Following regeneration, the ammonia-laden spent regenerant volume is about 2.5 percent to 25 percent of the throughput volume treated prior to regeneration, depending on the type and concentration of the regenerant.

Obviously, this regenerant cannot be disposed of to the aquatic environment.

The original approach to recovering and reusing the regenerant was to use a lime slurry as the regenerant so that the ammonium stripped from the bed during regeneration would be converted to gaseous ammonia which could then be removed from the regenerant by air stripping (17).

Regenerant with lime alone was found to be a rather slow process; therefore, the ionic strength of the regenerant solution was increased by the addition of salt (NaCl). The increased ionic strength of the regenerant plus the presence of sodium ion accelerates the removal of ammonia from the zeolite. Although most of the sodium chloride added to the regenerant is converted to calcium chloride by continuous recycle of the regenerant, sufficient sodium ion remains under steady-state conditions to promote the elution of the ammonium ions.

The use of the high pH regenerant is accompanied by an operational problem. Some plugging of the bed with $Mg(OH)_2$

and $CaCO_3$ occurs when the high pH regenerant is used. Attrition of the zeolite is aggravated by the violent backwashing needed to remove these solids.

In an alternate approach, ammonia in the regenerant solution is converted to nitrogen gas by reaction with chlorine which is generated electolytically from the chlorides already present in the regenerant solution. This process can be carried out with a regenerant of neutral pH so that the problem of precipitation of $Mg(OH)_2$ and $CaCO_3$ within the bed during regeneration is eliminated. Also cold weather does not affect this regenerant recovery process. The regenerant solutions used are rich in NaCl and $CaCl_2$ which provide the chlorine produced at the anode of the electrolysis cell. The reactions for the destruction of ammonia by chlorine are the same as for breakpoint chlorination.

During regeneration of the ion exchange bed, a large amount of calcium is eluted from the zeolite along with the ammonia. This calcium may be removed from the spent regenerant solution by a softening process prior to passing the spent regenerant through the electrolytic cells. The softening step would lower the calcium concentration below the level that would cause calcium hydroxide formation in the electrolytic cells. Unfortunately, return of the solids and the accompanying water resulting from this softening step to the plant solids handling system unavoidably results in an increase in effluent TDS. High flow velocities through the electrolysis cells are required in addition to a low concentration of calcium to minimize scaling of the cathode by calcium hydroxide and calcium carbonate. Acid flushing of the cells would be necessary to remove the scale when the cell resistance becomes too high for economical operation.

An alternate to air stripping or electrolysis of the regenerant is steam stripping. Stream stripping of regenerant is being practiced at the 0.6 mgd (U.S.) Rosemount, Minnesota physical-chemical plant (18). This process is economically feasible only with high pH regenerant. At Rosemount, ammonia is recovered from the spent ion exchange regenerant in an ammonia stripper. Steam is injected into a distillation column countercurrent with the regenerant solution to strip off the ammonia. An air-cooled, plate-and-tube condenser then condenses the vapor for collection in a covered tank as a 1 percent aqueous ammonia solution for sale as a fertilizer.

The steam stripping process is based on the use of high pH regenerant which has the disadvantages noted earlier. Battelle Northwest's evaluation of steam stripping indicates

that it is economically feasible if the regenerant volume is held to 4 bed volumes per cycle. A stripping tower depth of 24 feet and a loading of 7 gpm/sq ft are being used at Rosemount. Ceramic saddles are used rather than wooden slat packing because wood is not a suitable packing in a high pH steam environment.

Heat exchangers are used to transfer heat from the stripped regenerant to the incoming, cold regenerant. Heat transfer to the incoming regenerant from the condenser used to condense the stripped regenerant may also be attractive. Provisions for scale control in the heat exchangers should be provided. The steam requirements have been estimated to be 15 pounds per 1,000 gallons.

A new process for removal of ammonia and recovery of the regenerant was developed by CH2M HILL to eliminate many of the shortcomings of other methods of handling or recovery of spent regenerant from the clinoptilolite ion exchange process. This process was labeled ARRP (Ammonia Removal and Recovery Process).

The regenerant solution used with ARRP is sodium chloride. In the regeneration process sodium exchanges with ammonium on the clino resin. The regeneration cycle is a countercurrent process to minimize the makeup brine required, using four tanks of regenerant having different degrees of activity from virgin to spent.

The regenerant solution at equilibrium conditions contains approximately 2 percent sodium chloride, 1 percent calcium chloride, plus about 600 mg/l potassium chloride. Initially, the regenerant will be a 2 percent sodium chloride solution and this is the active regenerant. Calcium and potassium ions are allowed to build up in the regenerant solution until their concentration and the influent are in balance. No further significant removal of these ions occurs following this equilibrium. A small amount of magnesium will be present in the influent to the bed. This will be captured on the bed and subsequently removed from the bed during regeneration. Final removal of magnesium in the regenerant occurs by chemical precipitation following addition of sodium hydroxide.

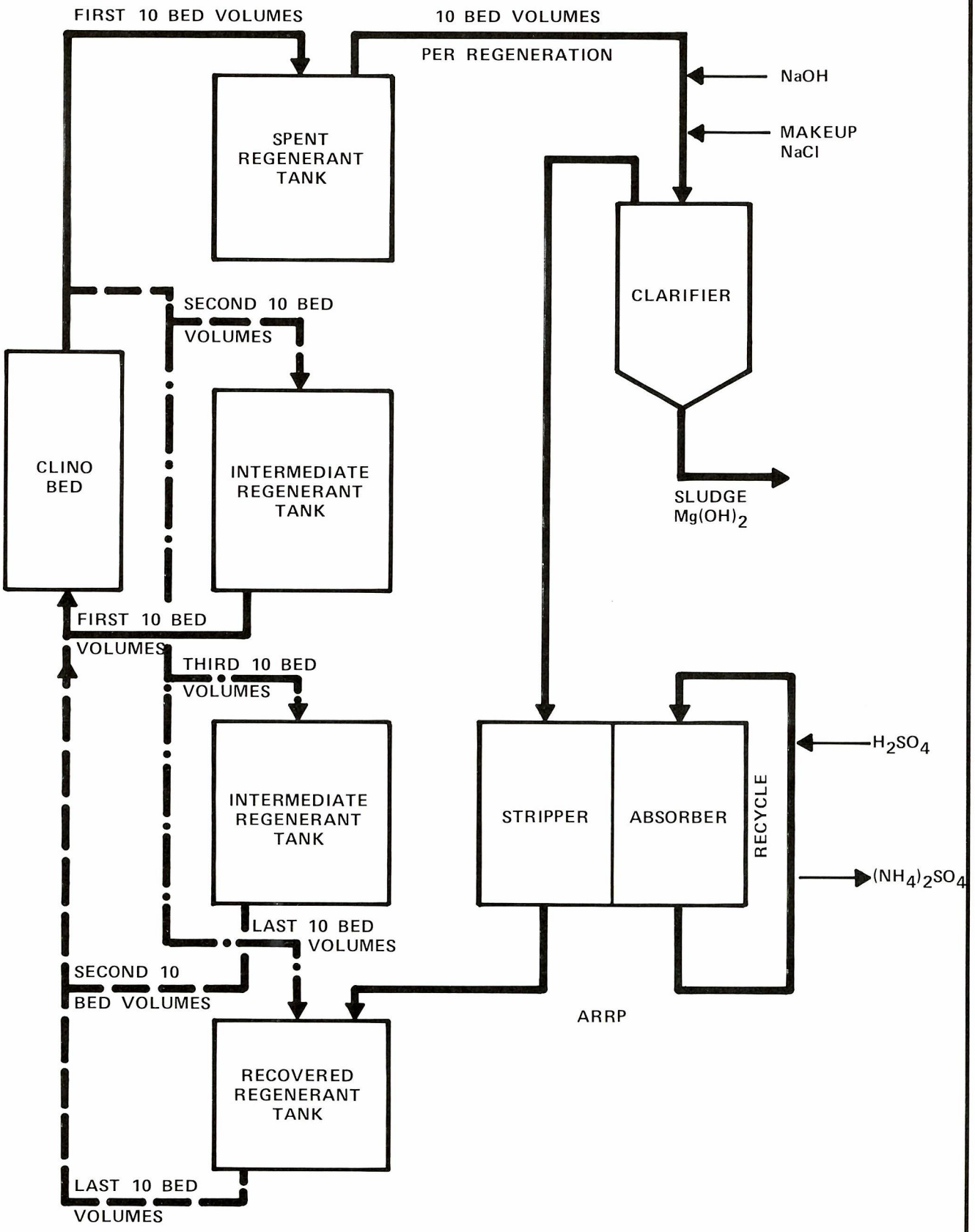
Spent regenerant stored in the tanks varies in ammonia nitrogen concentration from about 250 mg/l in the spent tank to about 50 mg/l in the recovered regenerant tank. The intermediate tanks have intermediate concentrations. This arrangement has been used to maximize the concentration in the spent tank while maintaining acceptable limits elsewhere in the system. A higher concentration obtained in the spent tank means that the ARRP and magnesium hydroxide

clarification facilities can be appropriately smaller. The regenerant pH varies from about 9.5 in the recovered regenerant tank to about 7.0 in the spent tank. Higher pH values produce more efficient regeneration due to the immediate conversion of ammonium ions to ammonia gas. However, this rather neutral pH level has been used to avoid problems with magnesium hydroxide precipitation in the bed during regeneration and also because of the fear of attrition of the clinoptilolite caused by high pH. If magnesium hydroxide precipitation occurs during regeneration, the regeneration cycle is plagued with a lingering high pH of about 10. Unless this high pH can be tolerated, considerable rinsing is required.

The normal regeneration sequence is shown on Figure 5. This simply involves pumping from one tank to the other through the ion exchange bed. A typical ammonia elution curve is shown on Figure 6 and the background concentrations in each tank is shown thereon. At the end of this cycle, the regenerant is recycled to the recovered tank. This has the effect of neutralizing the alkaline pH carryover from the ARRP process by converting the ammonium ions to ammonia gas. ARRP effluent is normally at a pH of 10.7 to 11.0. This is reduced to a pH of about 9.5 by recycling. In this manner, the pH is controlled without excessive use of acid and wasting of alkaline chemicals.

When spent regenerant is accumulated to a predetermined amount, the recovery portion of the process is activated. This system operates at a flow rate of approximately 7.7 percent of average AWT plant throughout, since the regenerant concentration is about 13 times as concentrated as plant waste. Initially, sodium hydroxide is added to achieve a pH of about 11. The amount added is stoichiometric for ammonium and magnesium ions removed. Sodium chloride is also added because of some salt loss from the regenerant solution in sludge removal and bed rinsing. Following pH adjustment, the regenerant is clarified and magnesium hydroxide is removed by sedimentation. Effluent is then pumped to the stripper-absorber columns for ammonia removal and recovery. A schematic of these process units is shown on Figure 7. The magnesium hydroxide sludge can be pumped to the inplant solids handling facility and mixed with other sludges for eventual disposal.

The ARRP process is a closed-air cycle, stripping and absorption system. The ammonia gas is first stripped into the air stream in a stripping tower. The air stream containing the ammonia is then directed into an absorption tower where the ammonia is absorbed. By maintaining the absorbent liquid in an acid condition, the absorbed ammonia is immediately converted to ammonium ions which are thus



■ FIGURE 5

SIMPLIFIED FLOW DIAGRAM OF REGENERATION SYSTEM USING ARRP PROCESS

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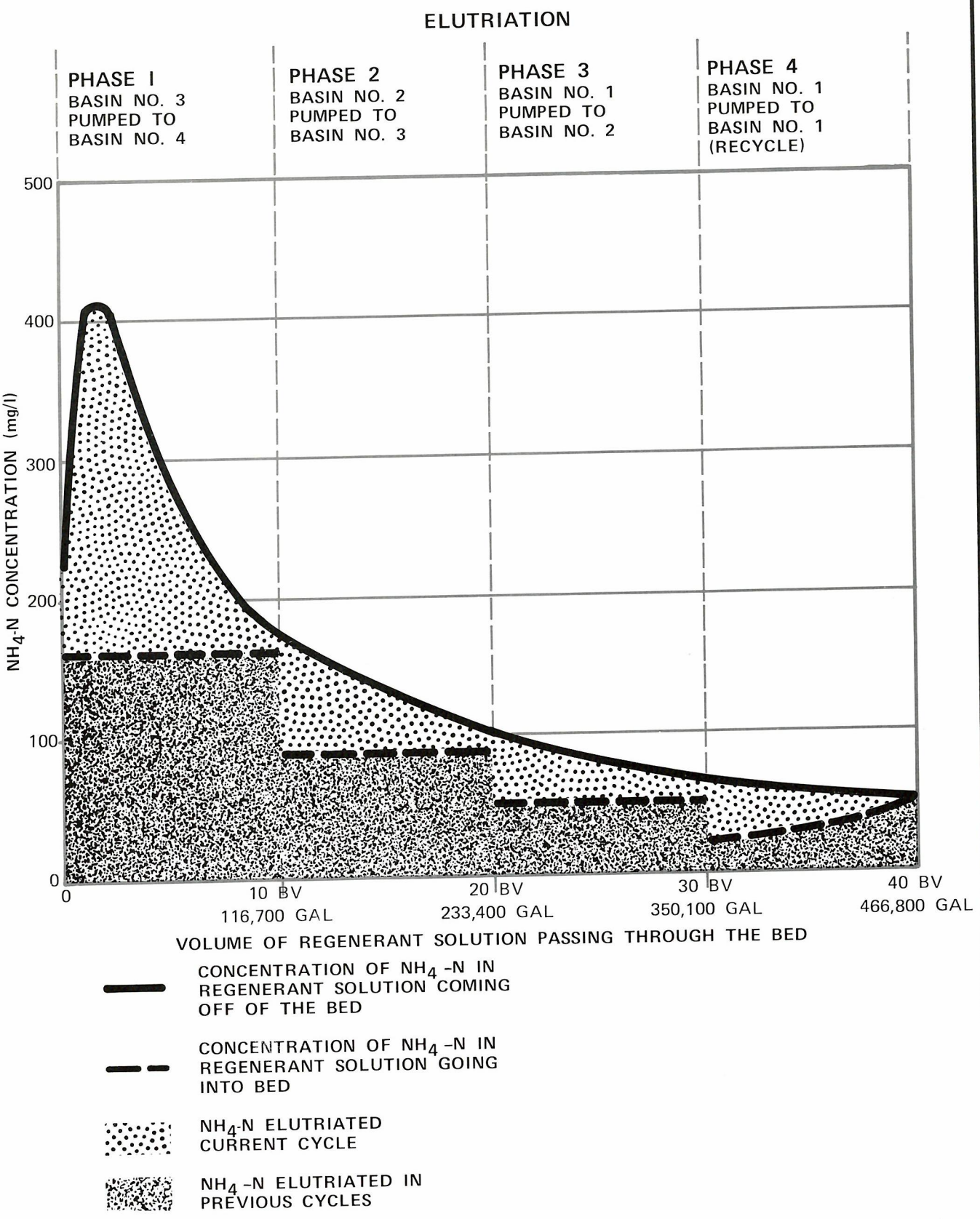


FIGURE 6

TYPICAL AMMONIA ELUTION CURVE

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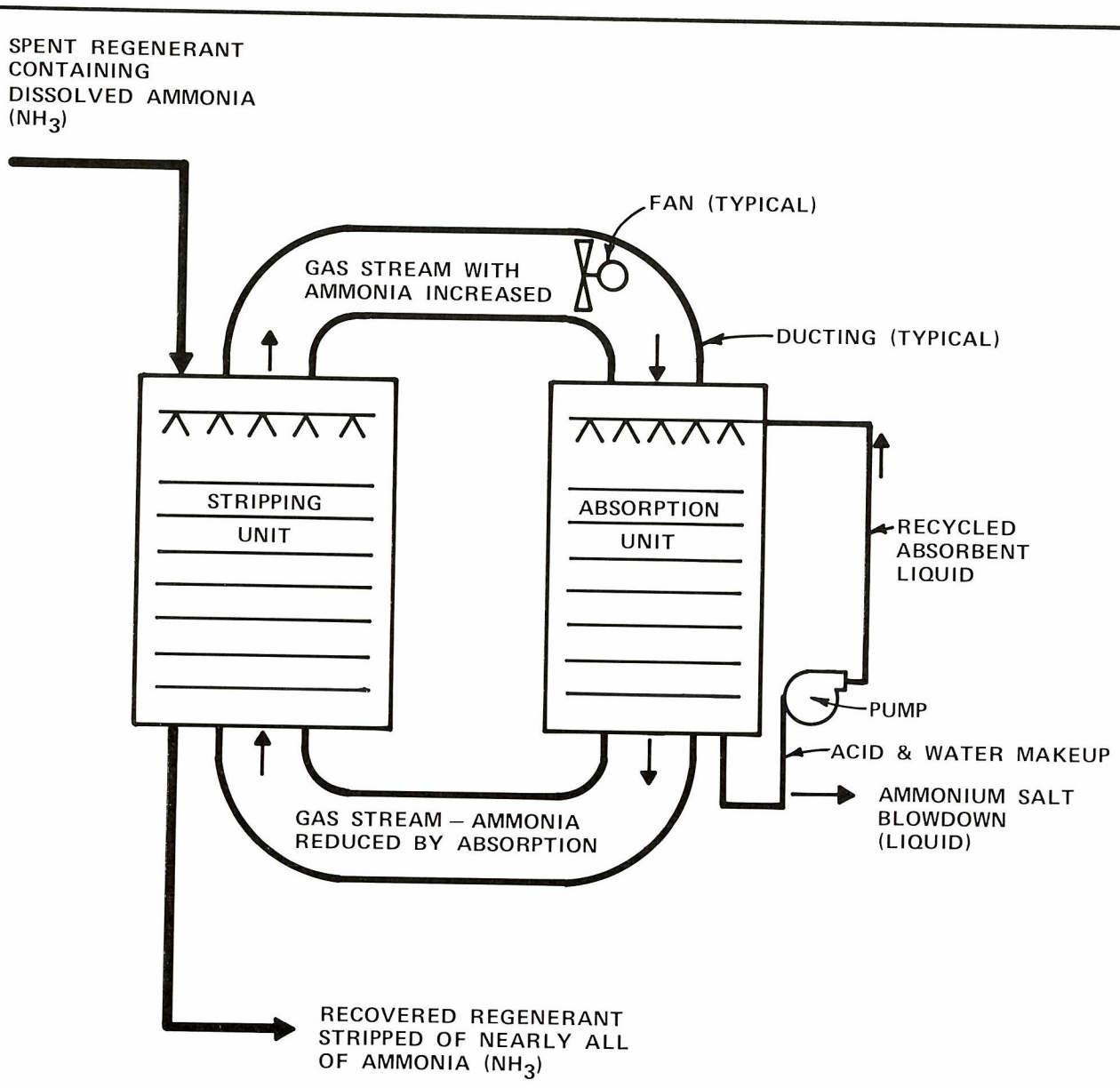


FIGURE 7

SCHEMATIC OF AMMONIA STRIPPING AND ABSORPTION UNIT

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CH2M HILL CANADA LTD

effectively trapped as the ammonium salt of the acid used. Using sulfuric acid, ammonium sulfate is formed. The concentration in the absorbent liquid is allowed to build up to approximately 40 percent ammonium sulfate and then blown down to a storage tank. Normally, the pH of the absorbent is maintained at about 2.0. It is neutralized by a small amount of sodium hydroxide addition when being pumped to the storage tank.

This process has operated very successfully at both the Upper Occoquan Sewage Authority's plant at Fairfax County, Virginia and at the Tahoe-Truckee Sanitary Authority's plant at Truckee, California. The process being entirely physical-chemical is extremely predictable and appears to have eliminated scaling problems associated with normal ammonia stripping caused by the presence of carbon dioxide in the air. It also eliminates problems of freezing and low efficiency at cold ambient temperatures. Further, it eliminates any fear of atmospheric or receiving water pollution caused by discharging ammonia gas into the air. The byproduct fertilizer has a wholesale value which is in excess of the cost of the acid addition. Depending on marketing conditions, the fertilizer sales (current U.S. pricing for ammonium sulfate is about $125 per ton) should significantly offset some of the other costs of this treatment process as well. In addition, the process is attractive from an energy standpoint in that the total energy required for production of the ammonium sulfate is only about two-thirds of what is normally required to produce ammonium sulfate commercially, including the energy required for offsite sodium hydroxide production.

Process Advantages

The advantages of selective ion exchange for ammonia nitrogen removal are: efficient ammonia removal in all seasons, and reliable efficiency of ammonia removal regardless of variation in influent ammonia concentration. Using the ARRP process for regenerant recovery provides a means of minimizing the amount of waste to be disposed of. Furthermore, a byproduct fertilizer is produced that has commercial value.

Process Disadvantages

Selective ion exchange accompanied by the regenerant recovery process is a more complex operation than other ammonia removal processes discussed in this paper. Many unit processes are included; these must be operated in concert with one another.

At Tahoe-Truckee, the calcium concentration of the influent to the clino beds has been found to be a significant consideration. Above 40 mg/l (as Ca), the calcium concentration reduces the bed volumes to exhaustion by a ratio of two bed volumes per mg/l of calcium. In some cases this may require the use of lime soda ash softening to maintain the service cycle of the clino beds between regenerations.

Because the clinoptilolite resin is a soft material, attrition of the resin due to backwashing and rinsing of the beds is significant. Scaling of the media is also a potential disadvantage that must be carefully considered in design. Present experience dictates that an acid wash be provided for both the resin and the stripping towers.

Furthermore, because ammonia is exchanged for sodium, there is a net increase in sodium, chloride, and TDS concentrations which may be of concern to water supplies downstream of the wastewater discharge. At the Tahoe-Truckee plant, experience to date has shown an increase in chloride and TDS of about 140 mg/l and 250 mg/l, respectively. This is largely due primarily to the clino/ARRP process.

System Design Considerations

Typical design criteria for the clino beds are as follows:

Volumetric loading: 6-10 bed volumes/hr. (BV/hr)
Surface loading: 4-6 gpm/s.f.
Bed Volumes to Exhaustion: 150-200
Backwash rate: 6 to 8 gpm/s.f.
Depth of media: 6 ft.
Influent calcium concentration: 40-50 mg/l

Typical design criteria for regeneration are:

Regenerant solution flow rate: 6.0 BV/hr.
Regenerant solution loading rate: 4.5 gpm/ft^2

Typical design criteria for the stripper and absorber towers in the regenerant recovery process are:

Hydraulic loading rate: 0.5 to 1.0 gpm/ft^2
Packing height: 9 to 10 ft.
pH to stripper: 11
pH of absorber solution: 2-3

Typical chemical usage at the Tahoe Truckee plant is as follows: (Based on infludent flow to the Clino beds.)

Sodium Chloride: 140 mg/l
Sodium Hydroxide: 60 mg/l
Sulfuric Acid: 95 mg/l

Other Physical Chemical Nitrogen Removal Processes

Stripping Ponds. The South Tahoe system has been modified to reduce the impact of temperature and scaling limitations encountered at the plant (18). Basically, the modified process consists of three steps: holding in high pH, surface agitated ponds; stripping in a modified, crossflow forced draft tower through water sprays installed in the tower; and breakpoint chlorination. This system was inspired by observations in Israel of ammonia nitrogen losses for high pH holding ponds.

Pilot tests at South Tahoe indicated that the release of ammonia from high pH ponds could be accelerated by agitation of the pond surface. In the modified Tahoe system, the high pH effluent from the lime clarification process flows to holding ponds with a detention time of 7-18 hours. The pond contents are agitated and recycled 4-13 times by pumping the pond contents through vertical spray nozzles into the air above the ponds. The holding pond detention time and number of recycles vary with plant flow, with the time and cycles decreasing as plant flow increases. At least 37 percent ammonia removal in the ponds is anticipated, even in cold weather conditions. The pond contents are then sprayed into the force draft tower. The packing has been removed from the tower and the entire area of the tower is equipped with water sprays. Based on pilot tests at least 42 percent removal of the ammonia in the pond effluent is anticipated. This added spray which includes recycling of the pond effluent through the tower to achieve 2-5 spraying cycles is effective even in cold weather. The ammonia escaping this process is removed by downstream breakpoint chlorination. It appears that stripping ponds offer an approach that takes advantage of the low cost and simplicity of the stripping process for removal of the bulk of the nitrogen, making breakpoint chlorination more attractive for remaining removal of ammonia.

Nitrate Ion Exchange. Research on nitrate removal by anion exchange has been conducted. This is due to interest by EPA and the water supply industry in removing nitrate-nitrogen at plants where drinking water limits are exceeded (19). Anion exchange using synthetic organic resin is less costly than reverse osmosis for removal of nitrate from water supplies. Disposal of the spent regenerant brine solution

containing nitrate is a potential problem. Another problem is the leaching of nonvolatile organic matter from all of the synthetic organic resins tested to date. Cost of this type of treatment is believed to be much higher than comparable process for ammonia removal.

Reverse Osmosis

Reverse osmosis will remove both ammonia and nitrate nitrogen from wastewater. The rejection efficiency of ammonia and nitrate will vary depending upon the flux rate used, the wastewater quality, and the type of membrane used. In general, use of reverse osmosis would not be practical for the sole purpose of nitrogen removal. However, if both TDS reduction and nitrogen removal is necessary, then reverse osmosis might be preferred over separate treatment processes for each.

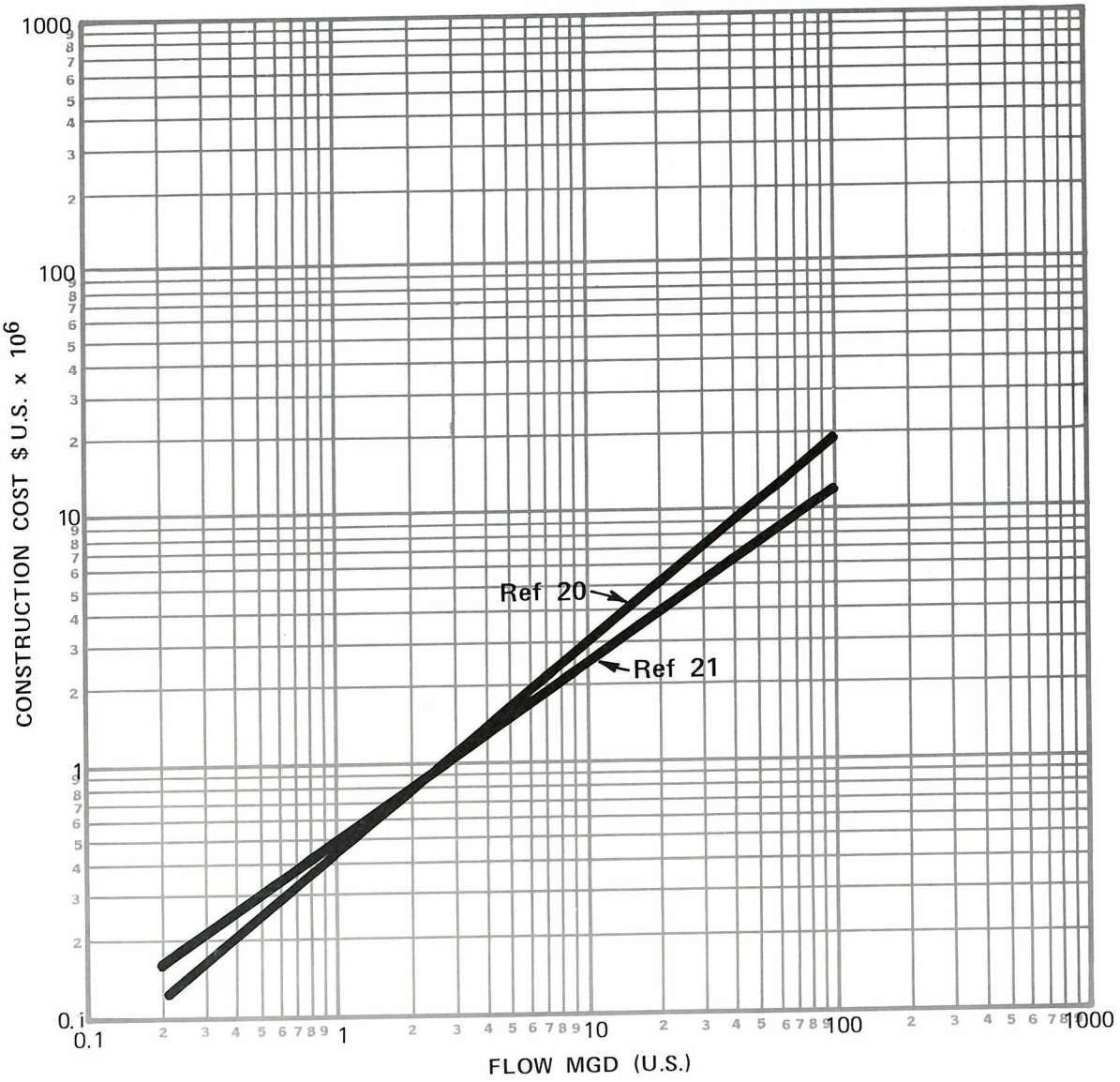
COST CONSIDERATIONS

Physical-chemical processes for the removal of nitrogen vary widely in construction cost as well as operation and maintenance cost. In addition, power requirements and use of chemicals also vary from process to process. Each method will therefore exhibit different sensitivity to changes in power, labour, and chemical costs.

Capital Costs

From the standpoint of capital costs, accurate projections are difficult because only a very lmited number of installations exist today and these are generally rather small in size and designed for quite specialized situations. Accurately projecting capital costs is further complicated by the fact that such data as exists does not correlate well between sources. This is in part due to varying design assumptions; different cost bases; and, most importantly, because the various sources have all excluded (or included) varying cost elements such as pumped vs. gravity flow, covered vs. uncovered basins, land, electrical and instrumentation or control loops.

Figures 8 through 11 present capital cost data from a number of sources for nitrogen removal by stripping, breakpoint chlorination, selective ion exchange and reverse osmosis. In all cases, the information presented has been adjusted from the basis used in the reference(s) (20, 21, 22, 23) to a uniform ENR 3000 index and an attempt has been made to standardize contingency, legal, fiscal, engineering, and construction interest costs at 35 percent. Land costs are excluded. The authors frankly have little faith in such



COST BASIS:

ENR INDEX 3000 – DOES NOT INCLUDE LIME CLARIFICATION OR RECARBONATION PROCESSES OR SLUDGE HANDLING AND DISPOSAL

HYDRAULIC LOADING –

Ref. 20 . . . . . .0.83 gal (Imp)/ft^2/min

Ref. 21 . . . . . .1.67 gas (Imp)/ft^2/min

AIR FLOW . . . . . 400 ft^3/gal

PUMP TDH . . . . . .50 ft

INFLUENT NH_4-N. . . .15 mg/l

EFFLUENT NH_4-N . . .2-3 mg/l

EQUATION FOR CURVES

COST $ U.S. (EPA) (20) = $ 523.640 $\times \left(\frac{\text{Capacity MGD Imp}}{0.83267}\right)^{0.79}$

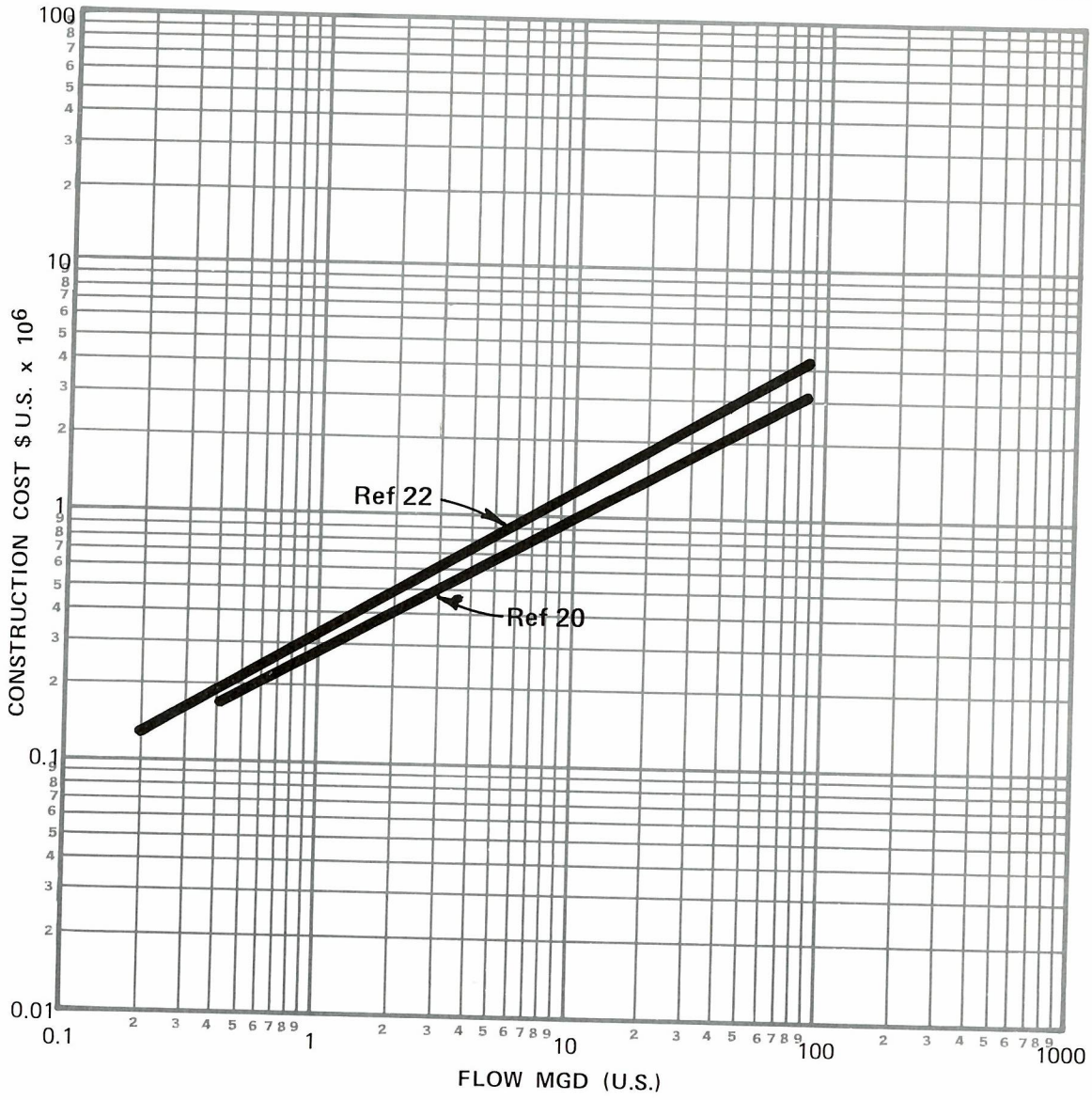
COST $ U.S. (CWC) (21) = $ 539.280 $\times \left(\frac{\text{Capacity MGD Imp}}{0.83267}\right)^{0.69}$

■ FIGURE 8

CONSTRUCTION COST FOR AMMONIA REMOVAL BY HIGH pH AMMONIA STRIPPING (ref. 20, 21)

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P181.A0



COST BASIS:

ENR INDEX 3000
GAS/SOLUTION Cl_2 FEED;
UNCOVERED TANKS NO OFF GAS STRIPPING;
COMPUTER CONTROL NOT INCLUDED;
NO DECHLORINATION

CONTACT TIME . . . .30 min

EQUATION FOR CURVES

COST $ U.S. (Bechtel) (22) $= 362.700 \times \left(\frac{\text{Capacity MGD Imp}}{0.83267}\right)^{0.54}$

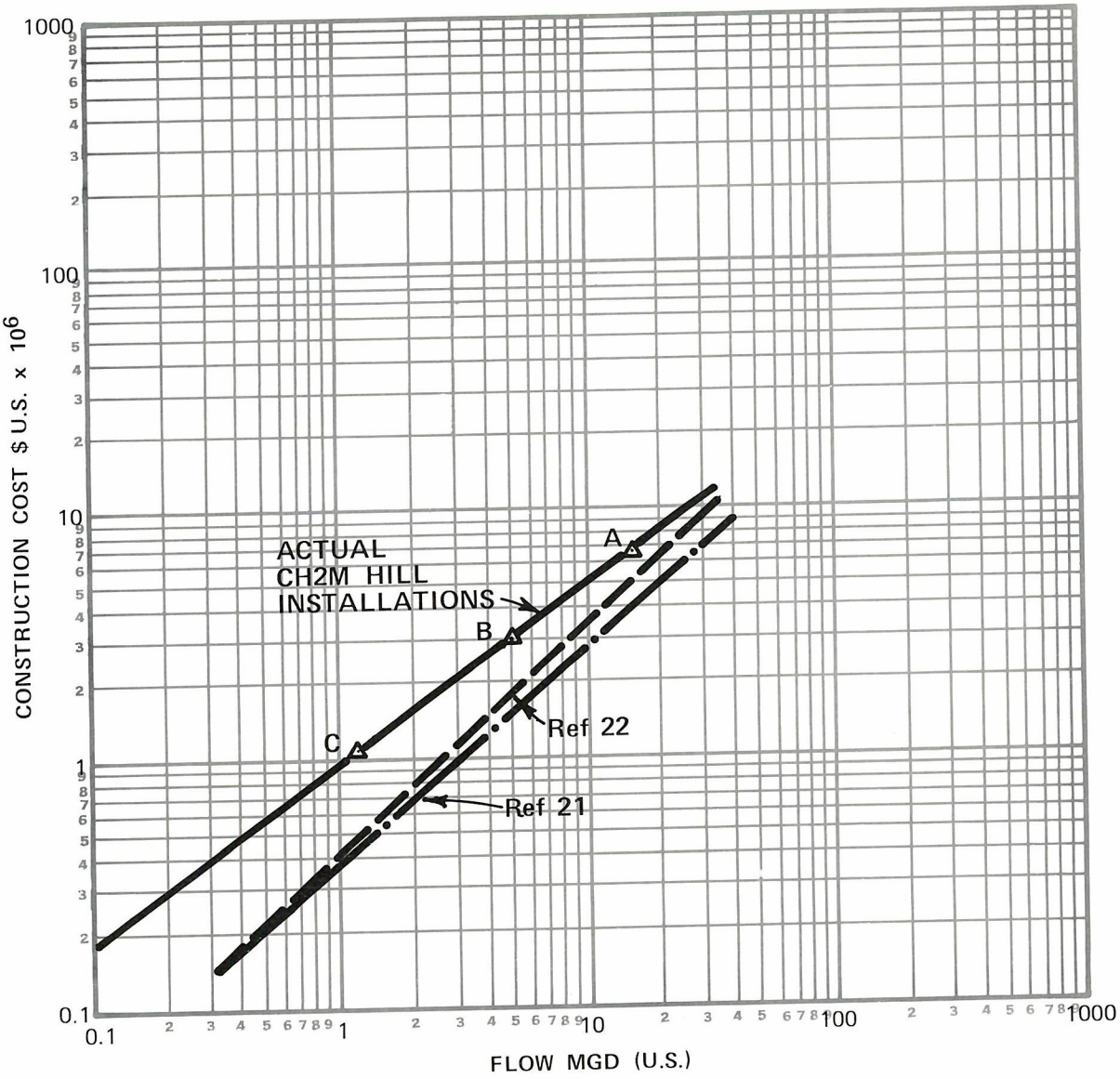
COST $ U.S. (EPA) (20) $= 327.000 \times \left(\frac{\text{Capacity MGD Imp}}{0.83267}\right)^{0.50}$

FIGURE 9

CONSTRUCTION COST FOR AMMONIA REMOVAL BY BREAKPOINT CHLORINATION (ref. 12, 21)

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COST BASIS

ENR INDEX = 3,000

EQUATION FOR CURVES

CH2M HILL $\text{COST \$ U.S.} = 925{,}500 \times \left(\frac{\text{Capacity MGD Imp}}{0.83267}\right)^{0.72}$

CWC $\text{COST \$ U.S.} = 422{,}160 \times \left(\frac{\text{Capacity MGD Imp}}{0.83267}\right)^{0.85}$

BECHTEL $\text{COST \$ U.S.} = 418{,}660 \times \left(\frac{\text{Capacity MGD Imp}}{0.83267}\right)^{0.93}$

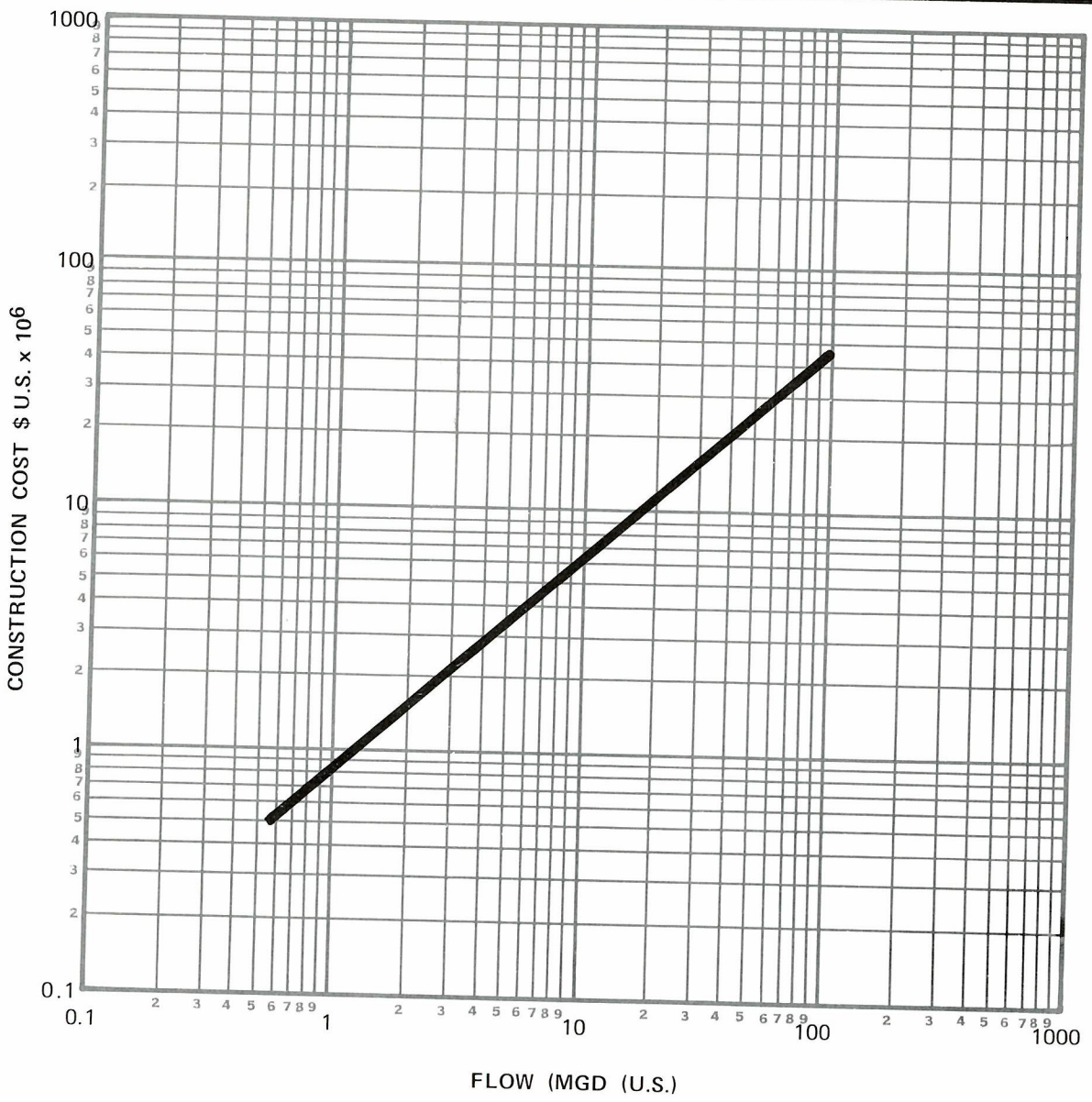
PLANT A = UPPER OCCOQUAN S.A.
B = TAHOE-TRUCKER S.A.
C = DENVER WATER BOARD (in design)

FIGURE 10

CONSTRUCTION COST FOR AMMONIA REMOVAL BY SELECTIVE ION EXCHANGE WITH REGENERANT RENEWAL BY CLOSED CYCLE STRIPPING (ref. 21, 22)

CH2M HILL CANADA LTD

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COST BASIS

ENR INDEX = 3,000

EQUATION FOR CURVE

$$\text{COST (\$ U.S.)} = 842{,}100 \left(\frac{\text{Capacity MGD-Imp}}{0.83267}\right)^{0.88}$$

FIGURE 11

CONSTRUCTION COST FOR REVERSE OSMOSIS (ref. 23)

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cost information and uniformly prefer to make project specific estimates. Recognizing however, that for comparative purposes, such data as exists is better than nothing, the cost curves thus derived are presented herein. The authors' opinion is that such cost curves are almost invariably low, usually by 30 percent or more. This seems to be well illustrated by the few actual data points plotted on the figures.

Operating and Maintenance Costs

Figures 12 through 15 present operation and maintenance cost data for nitrogen removal by stripping, breakpoint chlorination, selective ion exchange, and reverse osmosis. These data are taken directly from a 1978 publication (20) entitled "Innovative and Alternative Technology Assessment Manual," EPA 430/9-78-009. The basis fo costs used therein is listed in Table 2. Here also, the referenced costs, are in the authors' opinion, low. In particular, power usage figures shown are, for at least some unit processes, radically lower than actual operating experience would dictate. For this reason, Figures 16 through 20, taken from another EPA publication entitled "Energy Conservation in Municipal Wastewater Treatment" (23) are included herein. These portray annual electrical energy use as a function of capacity for the same four unit processes. The authors believe these latter data are far more realistic. By way of support for this opinion, consider a comparison of the reported values for power consumption for ammonia removal by selective ion exchange with regenerant renewal by closed cycle air stripping. From Figure 14, power usage for a 5 MGD (U.S.) ARRP system can be computed as $1,200 (U.S.) per year which at 2¢ per kwh amounts to 60,000 kwh/year or 12,000 kwh per year per MGD (U.S.) of flow. These costs are based on one regeneration per 24 hours and a clinoptilolite influent pumping head of 10 feet. On the other hand, Figure 18 indicates a power usage of 600,000 kwh per year or 120,000 kwh per year per MGD (U.S.) of flow for the regenerant renewal closed cycle air stripping process alone. Influent pumping at a TDH of 10 feet and regeneration pumping requirements would add an additional 29,000 kwh per year for a process total of about 149,000 kwh per MGD (U.S. per year. Actual measurements taken at the Tahoe-Truckee Sanitary Authority Treatment Plant indicate a total power usage of 200,000 kwh per year per MGD (U.S.) of plant flow for the regenerant renewal, influent pumping, and regenerant pumping.

TABLE 2

GENERAL COST BASIS FOR OPERATION AND MAINTENANCE COST CURVES

BASIS OF COSTS (All units U.S.)

LABOUR RATE, INCLUDING FRINGE BENEFITS = $7.50/hr

NOTE: LABOUR COSTS ARE BASED ON A MAN-YEAR OF 1,500 HOURS. THIS REPRESENTS: A 5-DAY WORK WEEK; AN AVERAGE OF 29 DAYS FOR HOLIDAYS, VACATIONS, AND SICK LEAVE; AND 6½ HOURS OF PRODUCTIVE WORK TIME PER DAY.

ENERGY COSTS

A. ELECTRIC POWER	=	$0.02/kWh
B. FUEL OIL	=	$0.37/gal
C. GASOLINE	=	$0.60/gal

LAND	=	$1,000/acre

CHEMICAL COSTS

A. LIQUID OXYGEN		=	$65/Ton
B. METHANOL		=	$0.50/gal
C. CHLORINE	150-lb CYLINDER	=	$360/ton
	1-ton CYLINDER	=	$260/ton
	TANK CAR	=	$160/ton
D. QUICKLIME		=	$25/ton
E. HYDRATED LIME		=	$30/ton (as CaO)
F. POLYMER (DRY)		=	$1.50/ton
G. FERRIC CHLORIDE		=	$100/ton
H. ALUM		=	$72/ton
I. ACTIVATED CARBON (GRANULATED)		=	$0.50/lb
J. SULFURIC ACID (66° Be)		=	$50/ton
K. SODIUM HEXAMETAPHOSPHATE		=	$0.25/lb
L. SO_2	150-lb CYLINDERS	=	$450/ton
	1-ton CYLINDER	=	$215/ton
	TANK CAR	=	$155/ton

DESIGN BASIS

OPERATION AND MAINTENANCE COSTS ARE BASED ON DESIGN AVERAGE FLOW.

OPERATION AND MAINTENANCE COSTS INCLUDE:

A. LABOUR COSTS FOR OPERATION, PREVENTIVE MAINTENANCE, AND MINOR REPAIRS.

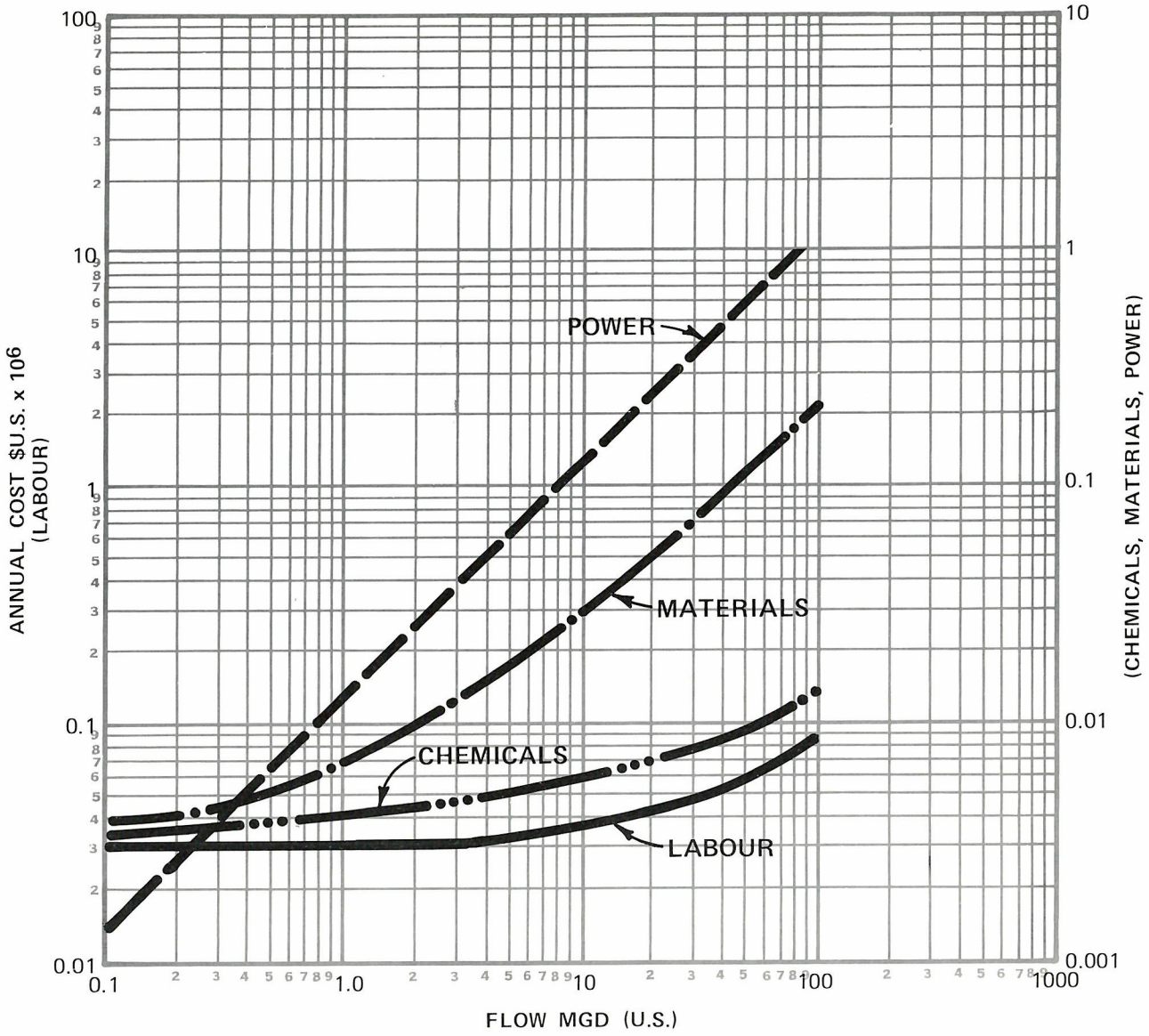
B. MATERIALS COSTS TO INCLUDE REPLACEMENT PARTS AND MAJOR REPAIR WORK (NORMALLY PERFORMED BY OUTSIDE CONTRACTORS).

C. CHEMICAL COSTS.

D. FUEL COSTS.

E. ELECTRICAL POWER COSTS.

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COST BASIS

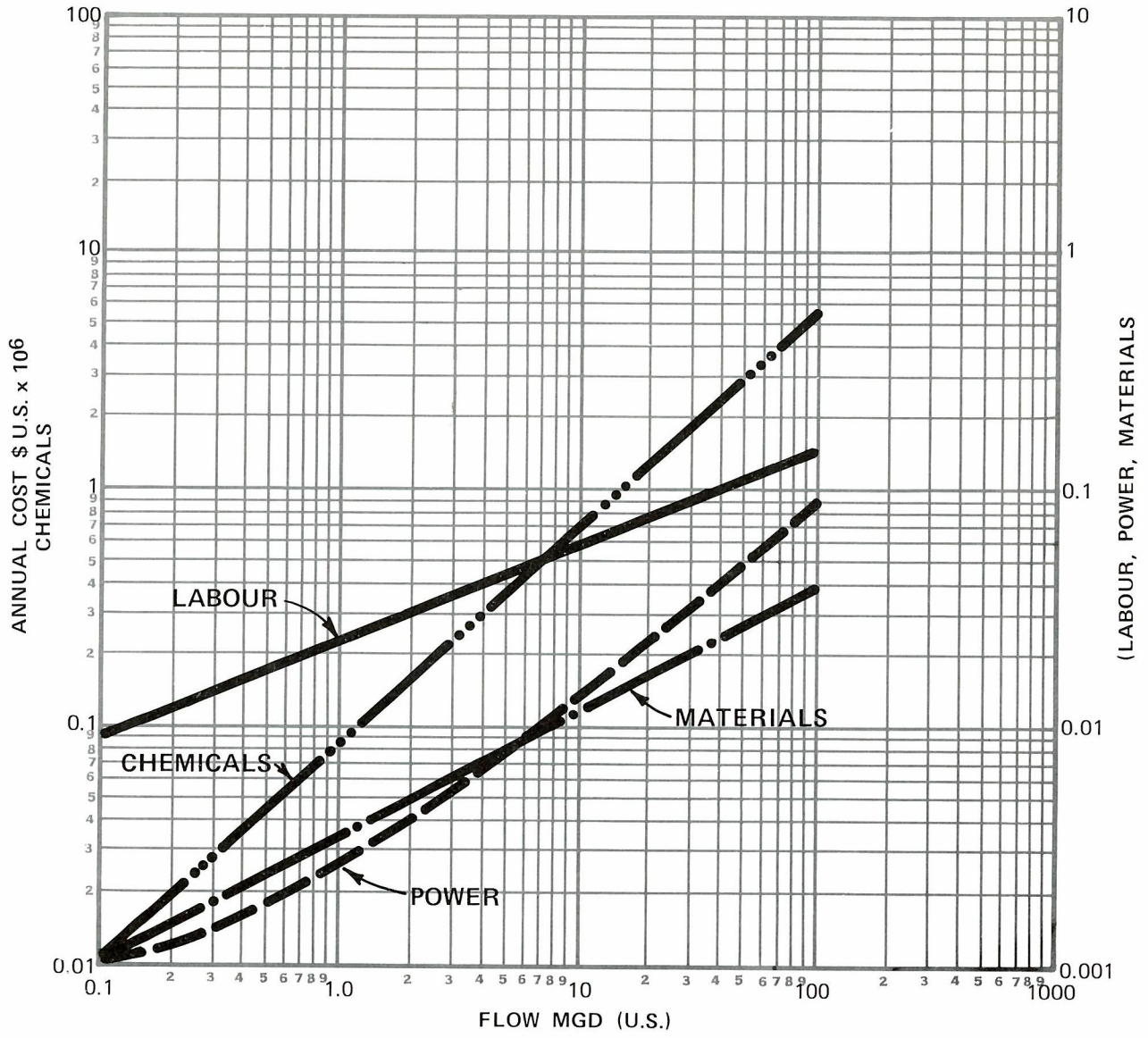
POWER. . . . . .$0.02 (U.S.)/kWh
LABOUR. . . . . .$7.50 (U.S.)/man hour
PUMP TDH . . . . . .50 ft
PUMP EFFICIENCY. . . .60% w/w
HYDRAULIC LOAD . . .1 gal (U.S.)/ft^2/min
AIR FLOW . . . . . .400 ft^3/gal (U.S.)
INFLUENT NH_3-N. . . . .18 mg/l
EFFLUENT NH_3-N . . . .3 mg/l

■ FIGURE 12

OPERATION AND MAINTENANCE COSTS FOR AMMONIA REMOVAL BY HIGH pH STRIPPING (ref. 20)

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COST BASIS

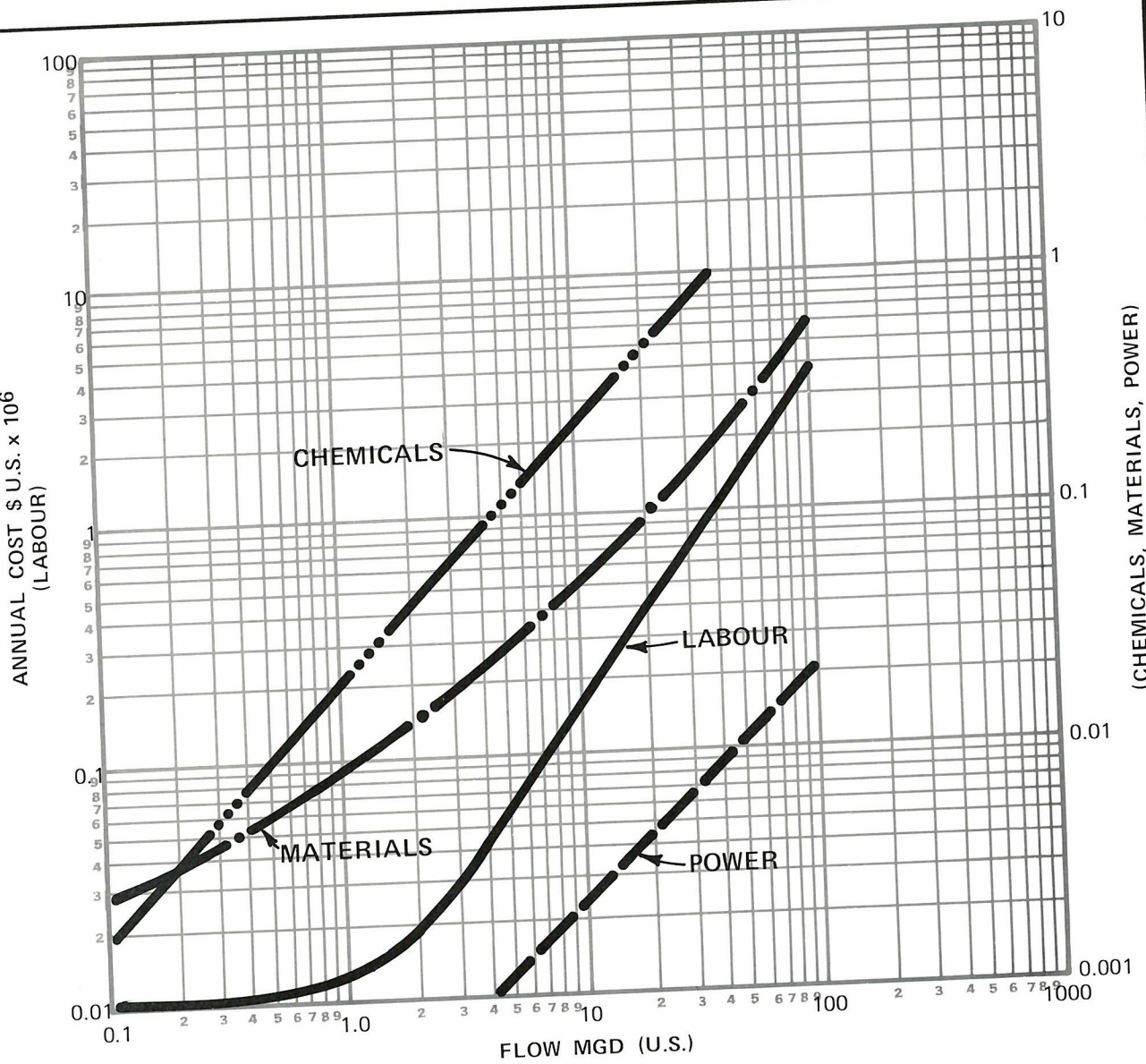
POWER . . . . . .$ 0.02 (U.S.)/kWh
LABOUR. . . . . .$ 7.50 (U.S.)/man-hour
LIME . . . . . .$ 160 (U.S.)/ton
NH_4-N INFLUENT. . . . .19 mg/l
NH-N EFFLUENT . . . . .2 mg/l
Cl_2 DOSE . . . . . 10 mg/mg NH_4-N
LIME DOSE . . . . . .0.9 mg/mg Cl_2

■ FIGURE 13

OPERATION AND MAINTENANCE COSTS FOR BREAKPOINT CHLORINATION (ref. 20)

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COST BASIS

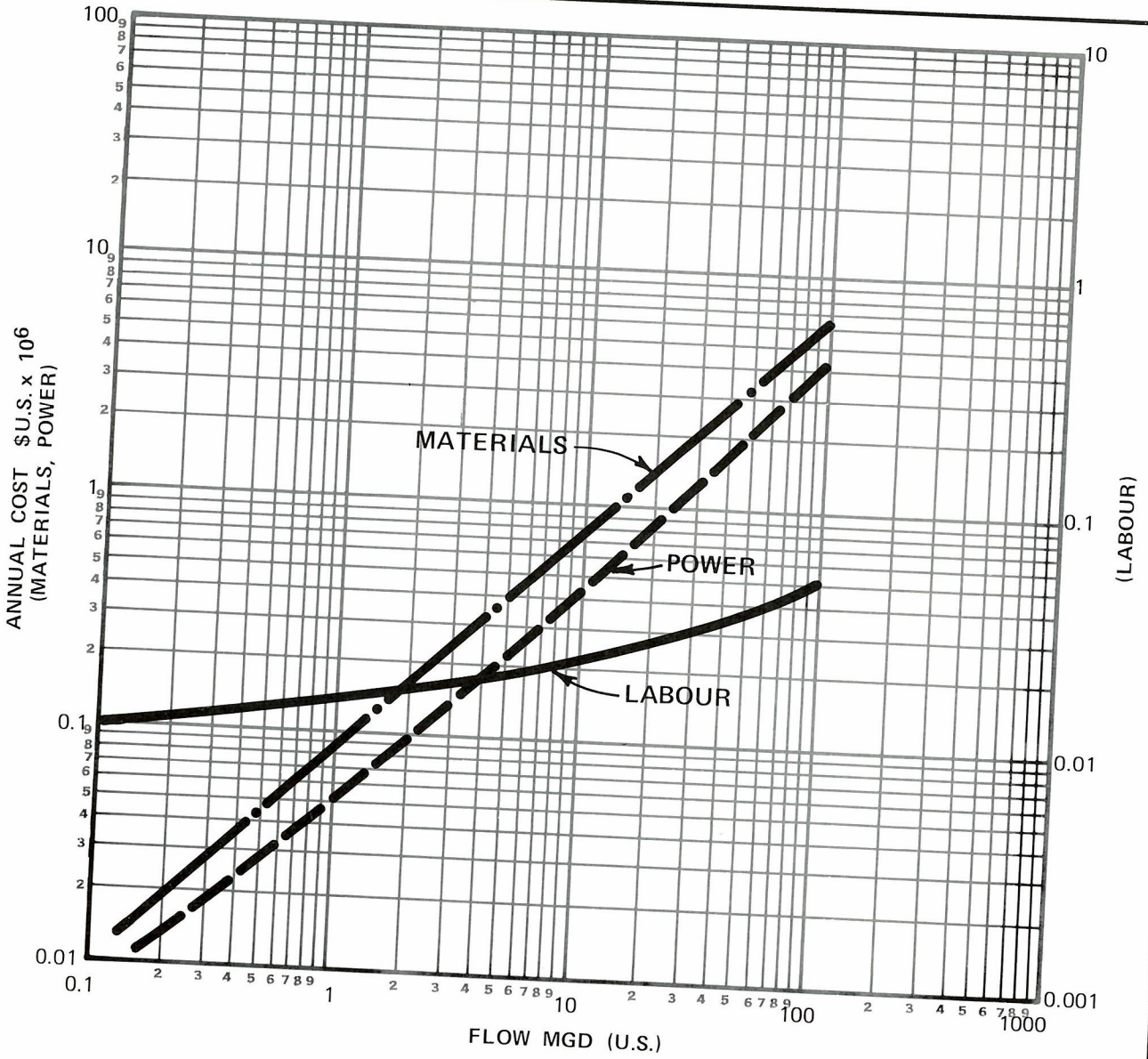
POWER. . . . . .$0.02 (U.S.)/kWh
LABOUR. . . . . .$7.50 (U.S.)/man hour
SULPHURIC ACID. .66° Be' $50 (U.S.)/ton
REGENERATION. . .40 bed Volumes/Cycle
THROUGHPUT . . . .100-150 bed Volumes/Cycle
REGENERANT . . . .2% NaCl Solution
SALT USE . . . . . .0.1 lb/1000 gals (U.S.) throughput
CAUSTIC SODA . . .0.9 mg NaOH/mg NH_4-N removed

■ FIGURE 14

OPERATION AND MAINTENANCE COSTS FOR AMMONIA REMOVAL BY SELECTIVE ION EXCHANGE WITH REGENERANT RENEWAL BY CLOSED CYCLE STRIPPING (ref. 20)

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COST BASIS

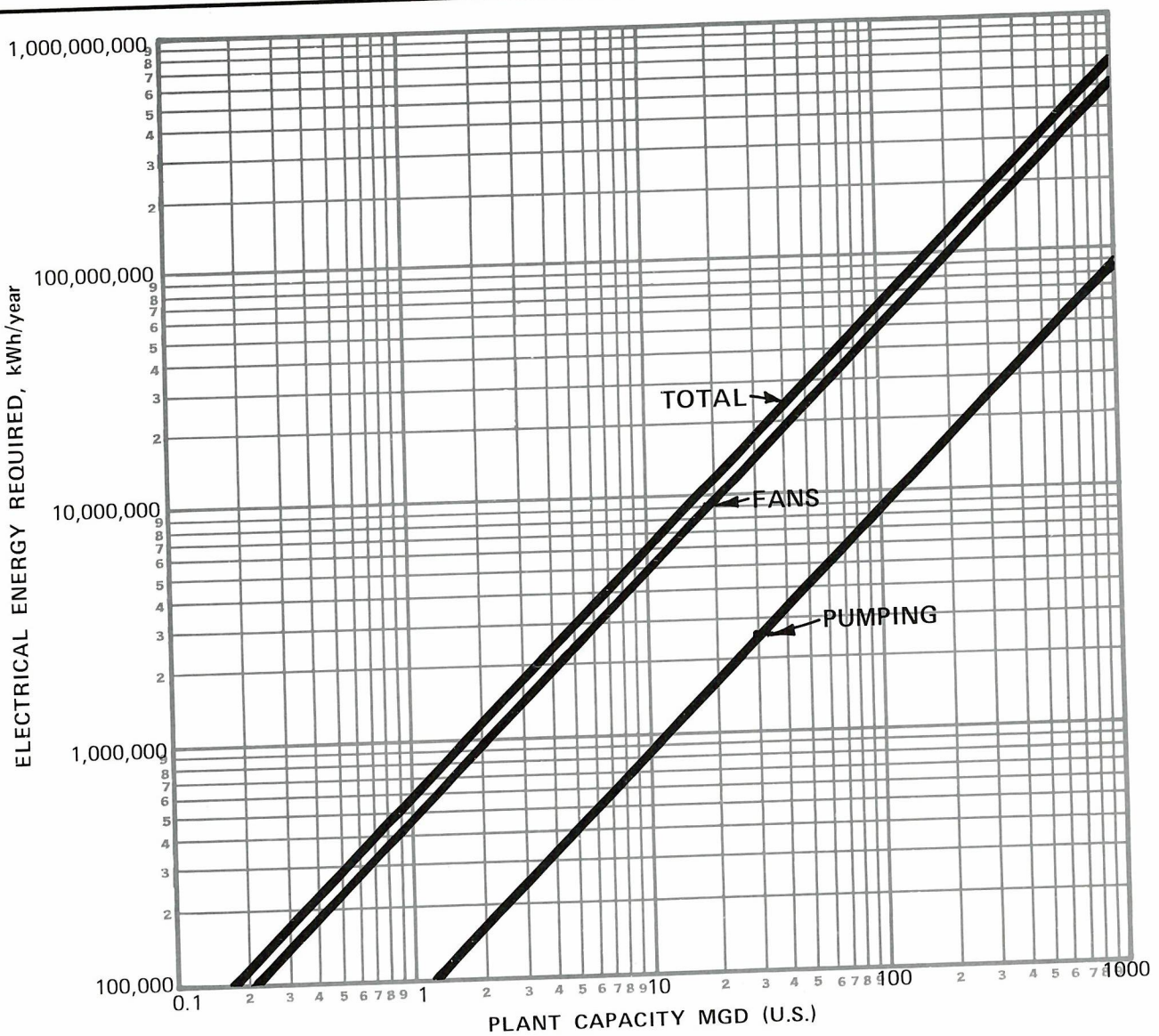
POWER. . . . . .$ 0.02 (U.S.)/kWh
LABOUR. . . . . .$ 7.50 (U.S.)/man-hour
OPERATION . . .@ 450 PSIG

■ FIGURE 15

OPERATION AND MAINTENANCE COSTS FOR REVERSE OSMOSIS. (ref. 20)

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DESIGN ASSUMPTIONS:

INFLUENT pH 11
AIR TEMPERATURE. . . . .21° C
INFLUENT NH_3-N. . . . . .15 mg/l
STRIPPING EFFICIENCY. .80%
HYDRAULIC LOADING. . .40.75 l/m^2/min
AIR/WATER RATIO . . . . .0.3 m^3/l

EQUATION FOR CURVES:

1. $kWh/year_{(pump)} = 83{,}333 \times \left(\frac{\text{Capacity (MGD Imp)}}{0.83267}\right)^{1.00}$

2. $kWh/year_{(fans)} = 500{,}000 \times \left(\frac{\text{Capacity (MGD Imp)}}{0.83267}\right)^{1.00}$

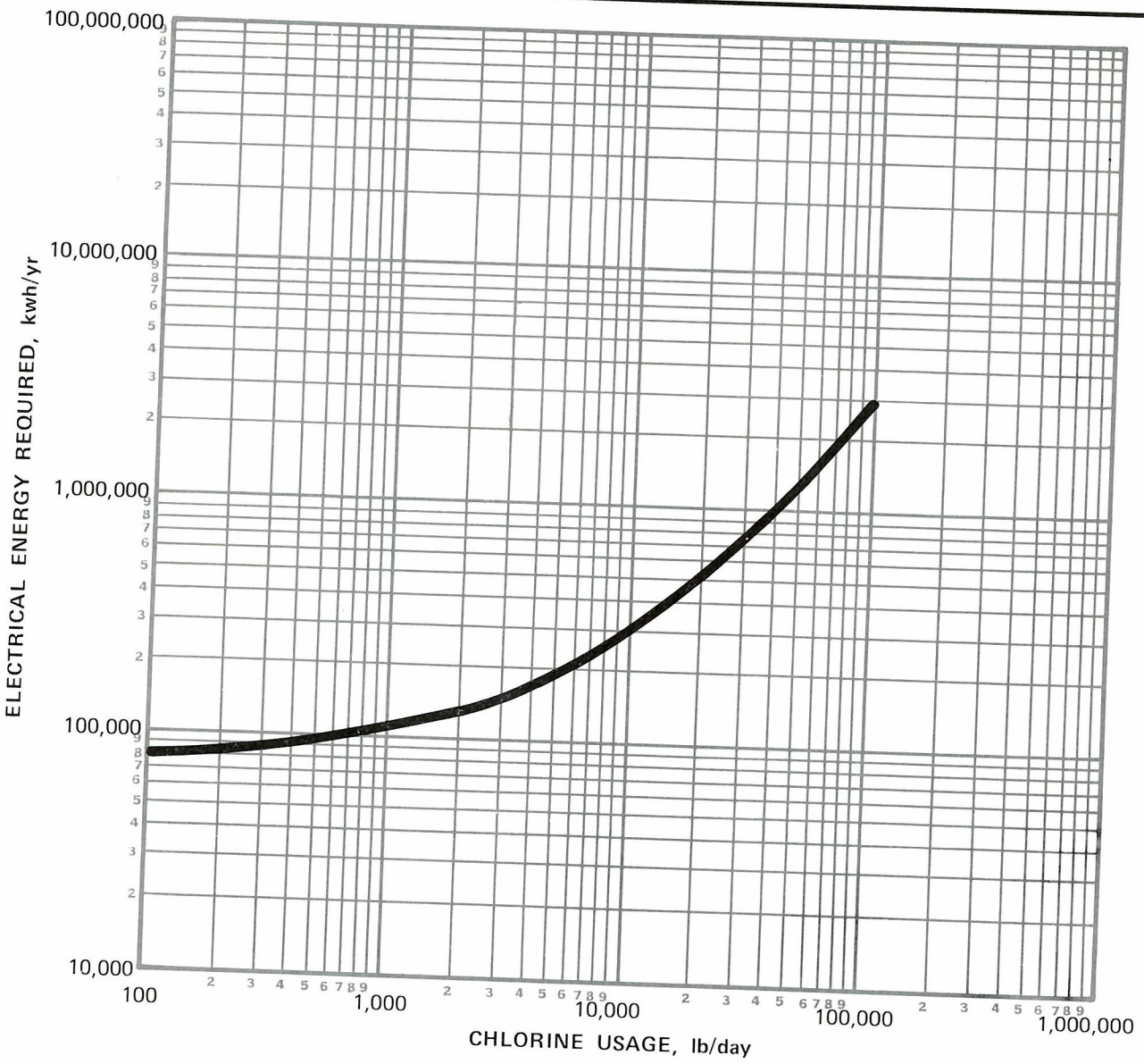
COEFFICIENT OF DETERMINATION (r^2) = 1.0

■ FIGURE 16

PRIMARY ENERGY REQUIREMENTS FOR HIGH pH AMMONIA STRIPPING (ref. 23)

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DESIGN ASSUMPTIONS:

INFLUENT NH_3-N	15 mg/l
EFFLUENT NH_3-N	0.1 mg/l
DOSAGE RATIO Cl_2:NH_3-N	8 : 1
RESIDUAL Cl_2	3 mg/l
DETENTION IN RAPID MIX	1 min
SULPHUR DIOXIDE FEED RATIO SO_2 : Cl_2	1 : 1

EQUATION FOR CURVE:

$$kWh/yr = 82{,}237 + 22.234(x) - 6.694 \times 10^{-5}(x^2) + 2.054 \times 10^{-9}(x^3)$$

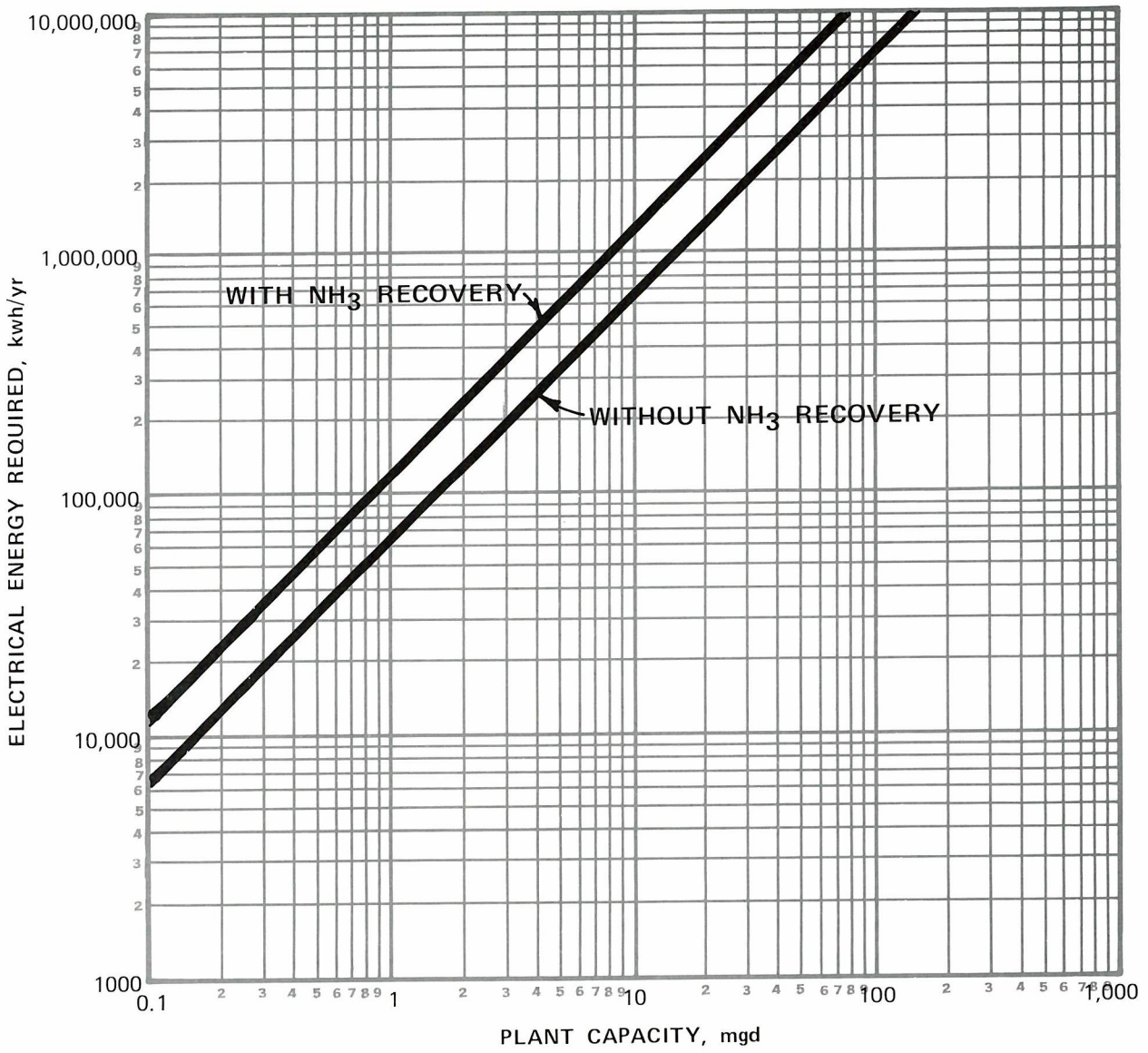
where x = lbs OF CHLORINE USED PER DAY

FIGURE 17

PRIMARY ENERGY REQUIREMENTS FOR BREAKPOINT CHLORINATION WITH SULPHUR DIOXIDE DECHLORINATION (ref. 23)

P181.A0

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DESIGN ASSUMPTIONS:

REGENERANT SOFTENED w/NaOH AND CLARIFIED AT 666 gal (Imp)/ft^2/day	
INFLUENT TO CLINO NH_3-N	15 mg/l
THROUGHPUT	150 Bed Volumes/Regeneration
REGENERATION	40 Bed Volumes Regeneration/Cycle
REGENERANT STRIPPER	633 gal(Imp)/ft^2
AIR FLOW	4.23 m^3 air/l
STRIPPER HEIGHT	32 ft

NH_3 RECOVERY BY ABSORBER TOWER WITH H_2SO_4

EQUATION FOR CURVES:

$$kWh/year_{(with\ recovery)} = 120{,}000 \times \left(\frac{CAPACITY\ MGD-IMP}{0.83267}\right)^{1.0}$$

$$kWh/year_{(without\ recovery)} = 6{,}667 \times \left(\frac{CAPACITY\ MGD/IMP}{0.83267}\right)^{1.0}$$

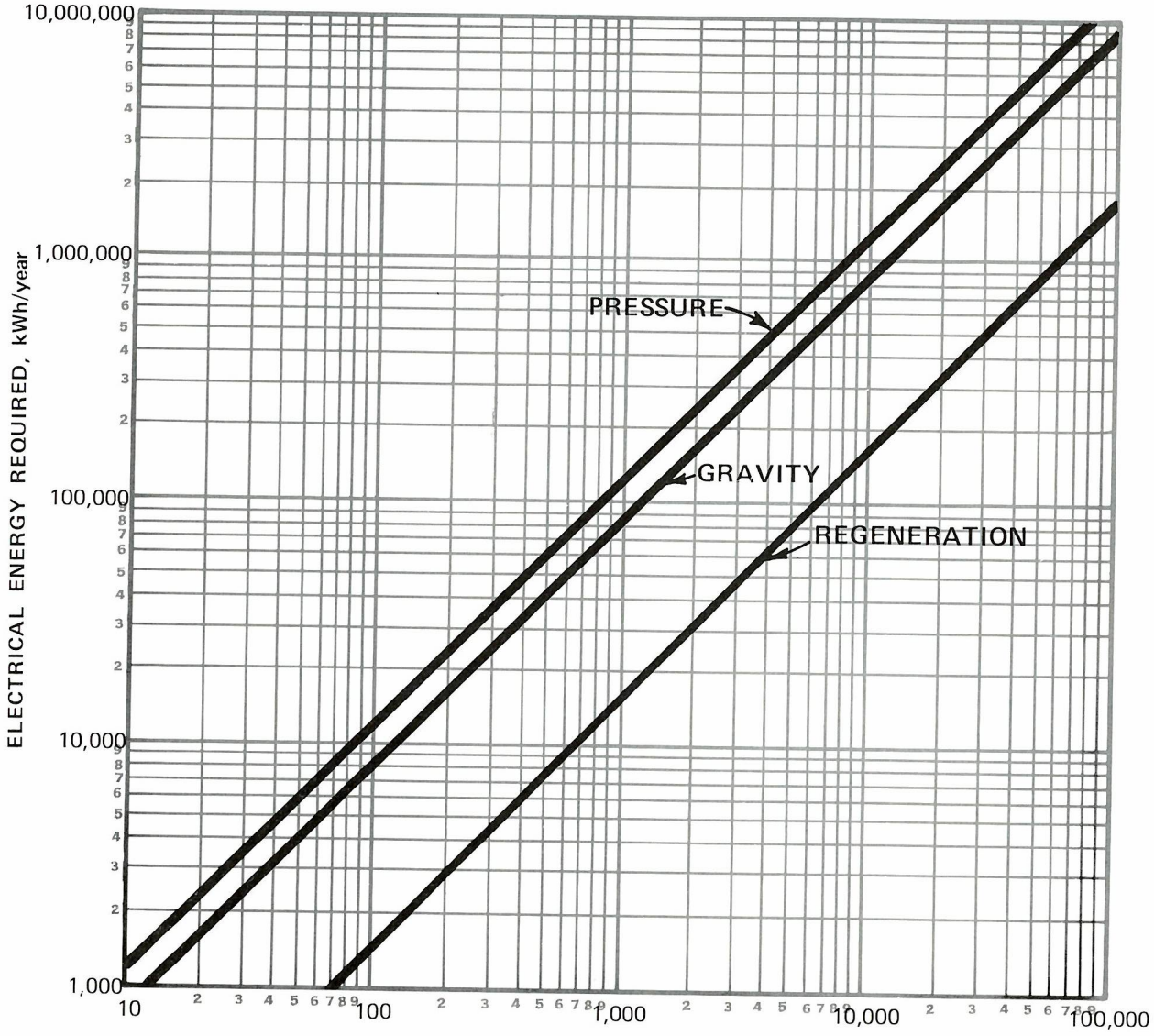
COEFFICIENT OF DETERMINATION (r^2) = 1.0

■ FIGURE 18

PRIMARY ENERGY REQUIREMENTS FOR AMMONIA REMOVAL BY ION EXCHANGE-REGENERANT RENEWEL BY AIR STRIPPING (ref. 23)

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DESIGN ASSUMPTIONS:

INFLUENT NH_3-N. . . . . .15 mg/l
INFLUENT SS. . . . . .$\leqslant$ 5.0
THROUGHPUT . . . . . .150 Bed Volumes
LOADING RATE. . . . . .6 Bed Volumes/hour
HEAD AVAIL TO GRAVITY . . . . .7.25 ft
PUMP HEAD FOR PRESSURE . . . .10 ft
REGENERATION. . . . . .40 Bed Volumes/Cycle
REGENERATION TIMING . . . . . .1 Regeneration/day
REGENERATION PUMP HEAD . . .10 ft

EQUATION FOR CURVES:

$kWh/year_{(gravity\ beds)}$ = 9.0 x (Bed Area - ft^2 @ 4 ft deep) $^{1.00}$ + Regeneration

$kWh/year_{(pressure\ beds)}$ = 125 x (Bed Area - ft^2 = 4 ft deep) $^{1.00}$ + Regeneration

$kWh/year_{(regeneration)}$ = 16.67 x (Bed Area - ft^2 @ 4 ft deep) $^{1.00}$

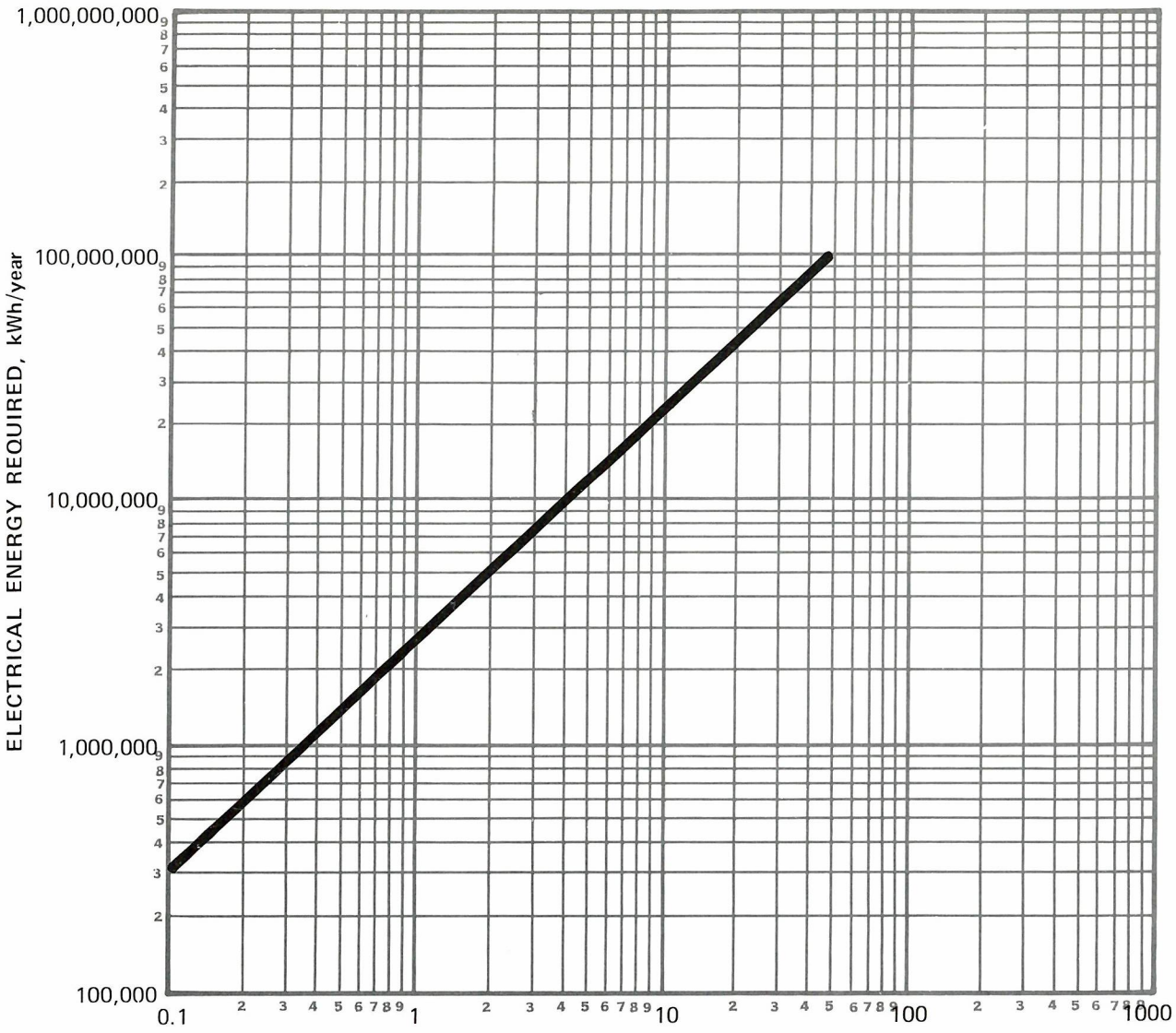
COEFFICIENT OF DETERMINATION (r^2) = 100

■ FIGURE 19

PRIMARY ENERGY REQUIREMENTS FOR AMMONIA REMOVAL BY ION EXCHANGE WITH REGENERATION (ref. 23) EXCLUDING REGENERANT RENEWAL

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DESIGN ASSUMPTIONS:

INFLUENT

- pH 6
- TURBIDITY . . . . . .< 1.0
- TDS . . . . . .≤ 1300
- NH_3-N 15
- NH_3-N REJECTION 85%
- FEED PRESSURE . . . . . .600 PSIG

EQUATION FOR CURVE:

$$\text{kWh/year} = 2.802 \times 10^6 \times \left(\frac{\text{Capacity MGD-Imp}}{0.83267}\right)^{0.95}$$

COEFFICIENT OF DETERMINATION (r^2) = 100

■ FIGURE 20

PRIMARY ENERGY REQUIREMENTS FOR AMMONIA REMOVAL BY REVERSE OSMOSIS (ref. 23)

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DESIGN EXAMPLE OF PHYSICAL CHEMICAL NITROGEN REMOVAL FOR NEW AND EXISTING PLANTS

Criteria

- Existing 20 MGD (Imp.) air activated sludge plant with anaerobic digestion and landfill disposal of vacuum filtered digested sludge; plant currently meets BOD and SS requirements. No nitrogen or phosphorus removal specifically provided; current effluent - 15 mg/l NH_3-N. Low effluent alkalinity.

- New standard to be met: <2.0 mg/l P and <4.0 mg/l total N and <2.0 mg/l unoxidized N year-round

- Location - Alberta Province - projected ENR index at midpoint of construction, 4,000.

Criteria Observations

- Location (temp.) and year-round N standard preclude air stripping without auxiliary heat.

- Phosphorous standard could be met much less expensively through iron salt addition or alum to activated sludge. This precludes high pH lime clarification which in turn precludes air stripping.

- Standard could be met by either selective ion exchange or breakpoint chlorination. No TDS standard imposed; therefore breakpoint chlorination may be feasible.

- Conclusion - Investigate selective ion exchange and breakpoint chlorination.

Capital Cost @ ENR 4000

- Breakpoint Chlorination

Cost - 362,700 x $\left(\frac{20}{0.83267}\right)^{0.54} \times \frac{4000}{3000} = \$2.691 \times 10^6$ (U.S.)
(Figure 9)

add computer controls and I&C @ 10% $\quad \frac{.269}{\$2.960} \times 10^6$

- Selective Ion Exchange

Cost 925,500 x $\left(\frac{20}{0.83267}\right)^{0.72} \times \frac{4000}{3000} = \underline{\$12.171} \times 10^6$ (U.S.)
(Figure 10)

Operating Cost

- Cost Basis:

Power	$0.05/kWh
Labour	$12.00/manhr (w/fringes)
Cl_2	$320/ton
Lime	$60/ton
Salt	$40/ton
NaOH	$240/ton
H_2SO_4	- $75/ton

- Breakpoint (assume 15 mg/l NH_3-N)

Cl_2 9 mg/l Cl_2/mg/l NH_3-N

NaOH 1.3 mg/l NaOH/mg/l Cl_2

Cl_2 $9 \times 15 \times \frac{20}{0.83267} \times 8.34 \times 365 \times \frac{1}{2000} \times \$320 =$

$\$1.579 \times 10^6$ (U.S.)

NaOH $9 \times 15 \times \frac{20}{0.83267} \times 1.3 \times 8.34 \times 365 \times \frac{1}{2000} \times 240 = \$1.540 \times 10^6$

Power 800,000 kWh x 0.05 = $\$ .040 \times 10^6$

Labour $\$ .058 \times 10^6$

$\$3.217 \times 10^6$/year

- Selective Ion Exchange (assume 15 mg/l NH_3-N)

Salt (makeup) $140 \times \frac{20}{0.83267} \times 8.34 \times 365 \times \frac{1}{2000} \times \$40 =$ $\$0.205 \times 10^6$

H_2SO_4 $95 \times \frac{20}{0.83267} \times 8.34 \times 365 \times \frac{1}{2000} \times \$75 =$ $\$0.260 \times 10^6$

NaOH $60 \times \frac{20}{0.83267} \times 8.34 \times 365 \times \frac{1}{2000} \times \$240 =$ $\$0.526 \times 10^6$

Power	3,671,000 kWh x 0.05	$0.183 x 10^6
Labour		$0.800 x 10^6
Clino (makeup) 70t/year @ $350/ton		$0.024
		$1.998 x 10^6/year

Total Annual Cost

- Cost Basis
  (20 yr @ 6-7/8%)

- Breakpoint

  Capital $2.960 \times 10^6 \times .09348 =$ $277,000/yr
  O&M 3,217,000/yr
  $3,494,000

  Cost per 1000 gals (Imp.)
  $$\frac{3,494,000}{20 \times 365 \times 1000} = \$0.48$$

- Selective Ion Exchange

  Capital $12.171 \times 10^6 \times .09348 =$ $1,138,000
  O&M 1,998,000
  $3,136,000

  Cost per 1000 gals (Imp.)
  $$\frac{3,136,000}{20 \times 365 \times 1000} = \$0.43$$

Preliminary Clinoptilolite Bed Design

Volumetric Loading - 6 to 10 B.V./hr

req'd volume of Clino bed $= 20 \times \frac{1}{24} \times \frac{1}{6.23} \times \frac{1}{6} \times 10^6 = 22,290 \text{ ft}^3$

Check surface load @ bed depth = 6':

$$\frac{22,293 \text{ ft}^3}{6 \text{ ft}} = 3715 \text{ ft}^3$$

$$20 \times \frac{1}{1440} \times 10^6 = 13,890 \text{ gpm}$$

$$13,890 \times \frac{1}{3715} = 3.74 \text{ gal/ft}^2\text{/min} \quad (4.49 \text{ gpm/ft}^2 \text{ U.S.})$$

Throughput to exhaustion:

@ 150 B.V. = 22,290 x 150 x 6.23 = 20.83 MG
@ 200 B.V. = $\frac{200}{150}$ x 20.83 = 27.77 MG

Regeneration Requirements

40 B.V. @ 6 B.V./hr = 6.67 hrs. + rinse time
say 8 hrs total

Therefore, a bed service cycle is 24 hours throughput + 8 hrs regeneration time and actual use cycle is 24/32 = 75% and total bed volume required is 22,290/.75 = 29,700 ft^3

Stripper Requirements

Spent regenerant per 24 hrs is 10 B.V. or 222,900 ft^3

222,900 x 6.23 = 1,389,000 gal/day or 964 gal/min

Stripper area @ 0.63 gal/min:

$$964 \times \frac{1}{.63} = 1530 \text{ ft}^2$$

Determine number of stripper modules:

@ 12'Ø (113 ft^2) $\frac{1530}{113}$ = 14 + spares

Size Individual Components

Regeneration Cycle:

Phase 1 - Regen from Tank 3 through bed to Tank #4 - 2 hrs
Phase 2 - Regen from Tank 2 through bed to Tank #3 - 2 hrs
Phase 3 - Regen from Tank 1 through bed to Tank #4 - 2 hrs

Each regeneration exhausts 10 B.V. of regenerant. If beds are sized for 4 in service, then at worst condition, 2 beds are in regeneration. Unless stripper system is designed for worst condition, i.e., 20 B.V.'s in 2 hrs, regenerant storage will be required. Storage is more economical.

Try 6 beds; 16 strippers; 2 sets tankage

Then bed size = $\frac{29{,}700 \text{ ft}^3}{6 \text{ ft}} \times \frac{1}{6 \text{ beds}}$ = 825 ft^2 for each of 6

Storage req'd total = 20 B.V. x 4 tanks = 80 B.V.
80 x 825 = 66,000 ft^3

w/2 sets of 4 tanks each; each tank is:

$$\frac{66,000}{8} = 8,250 \text{ ft}^3$$

With this configuration, the spent regenerant tanks have a maximum flow input during a 2 hour peak condition of 20 B.V. or 16,500 ft^3 - simultaneously the outflow to the stripper is

$$\frac{40 \text{ B.V.}}{24 \text{ hrs}} \times 2 \text{ hrs} = 3.33 \text{ B.V.}$$

Therefore a 20 B.V. tankage size will provide sufficient stroage to allow essentially average flow conditions of spent regenerant to the air stripping system.

Obviously, the foregoing illustrative design example is extremely brief. It is intended merely as an illustration of only a few of the myriad of considerations required in a real design. Certainly other configurations based on site constraints, periodic and diurnal flow fluctuations, reliability and redundancy requirements as well as a host of other factors would be investigated during an actual study and design.

References

1. Gonzales, John G. and Culp, Russell L. "New Developments in Ammonia Stripping." Public Works. May 1973.

2. Metcalf and Eddy, Inc. Wastewater Engineering. McGraw-Hill. 1972.

3. Water Pollution Control Federation. Wastewater Treatment Plant Design (MPO-8). 1977.

4. Eckenfelder, W.W, Manual of Treatment Processes. Vol. 1. Environmental Science Services Corporation. 1968.

5. Masselli, J.W., et al. "The Effect of Industrial Wastes on Sewage Treatment." New England Interstate Water Pollution Control Commission. June 1965.

6. McCarty, P.L. et al. "Water Factory 21: Reclaimed Water, Volatile Organics, Virus, and Treatment Performance." U.S. Environmental Protection Agency Publication EPA-600/2-78-076. June 1978.

7. Miner S. "Preliminary Air Pollution Survey of Ammonia." U.S. Public Health Service. Contract No. PH22-068-25. October 1969.

8. Culp, Russell L., et al. Handbook of Advanced Wastewater Treatment. 2nd edition. Van Nostrand Reinhold Company. 1977.

9. U.S. Environmental Protection Agency. Technology Transfer. Process Design Manual for Nitrogen Control. October 1975.

10. Wesner, G.M. and Culp, R.L. "Wastewater Reclamation and Seawater Desalination." JWPCF. Vol. 44, p. 1932. October 1972.

11. Dean, R.B., ed. "Nitrogen Removal from Wastewaters." FWQA, Division of Research and Develoment, Advanced Waste Treatment Research Laboratory. Cincinnati, Ohio. May 1970.

12. O'Farrell, T.P., et al. "Nitrogen Removal by Ammonia Stripping." JWPCF. 44:8, p. 1527. August 1972.

13. U.S. Environmental Protection Agency. Technology Transfer. "Physical-Chemical Nitrogen Removal--Wastewater Treatment." July 1974.

14. Pressley, T.A., et al. "Ammonia Removal by Breakpoint Chlorination." Environmental Science and Technology." Vol. 6, No. 7, p. 622. July 1972.

15. Van Vuuren, L.R.J., et al. "Stander Water Reclamation Plant: Chlorination Unit Process." Project Report 21. Pretoria, So. Africa. November 1972.

16. E.S. Barrekette, ed. "Pollution: Engineering and Scientific Solutions" Plenum Publishing Corp. New York. p. 522.

17. Battelle-Northwest, "Ammonia Removal from Agricultural Runoff and Secondary Effluents by Selective Ion Exchange." Robert A. Taft Water Research Center Report No. TWRC-5. March 1969.

18. U.S. Environmental Protection Agency. Technology Transfer. Process Design Manual for Nitrogen Control. October 1975.

19. Clifford, Dennis A. and Weher, Walter J. "Nitrate Removal from Water by Ion Exchange." U.S. Environmental Protection Agency Report No. 600/8-77-015. November 1977.

20. EPA (Burns and Roe Consulting Engineers). Innovative and Alternative Technology Assessment Manual. U.S. Environmental Protection Agency Report No. 430/9-78-009. 1978.

21. CWC Consulting Engineers. "Estimating the Costs of Wastewater Treatment Facilities. Virginia State Water Control Board Report. March 1974.

22. Bechtel Inc. A Guide to the Selection of Cost Effective Wastewater Treatment Systems. U.S. Environmental Protection Agency Report. No. 430/9-75-002. July 1975.

23. CWC Consulting Engineers. Estimating Costs for Water Treatment Plant Efficiency. U.S. Environmental Protection Agency Report No. 600/2-78-182. August 1978.

24. CWC Consulting Engineers. Energy Conservation in Municipal Wastewater Treatment. U.S. Environmental Protection Agency Report No. 430/9-77-011. March 1977.

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APPENDIX
Reproductions of Slides used in Presentation

ESTIMATED ENERGY REQUIREMENTS FOR THE PRODUCTION OF CONSUMABLE CHEMICALS

MATERIAL	FUEL (Million Btu/ton)	ELECTRICITY (kWh/ton)
CARBON DIOXIDE	2	200
CHLORINE	42	4,000
CALCIUM OXIDE (LIME)	5.5	600
SODIUM HYDROXIDE	37	36,000
SULPHUR DIOXIDE	0.5	48
SULPHURIC ACID	1.5	200

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REACTIONS OF CHLORINE AND AMMONIA

1) $NH_4^+ + HOCl \rightarrow NH_2Cl$ (monochloramine) $+ H_2O + H^+$

2) $2NH_2Cl + HOCl \rightarrow N_2 + 3HCl + H_2O$

3) $NH_2Cl + HOCl \rightarrow NHCl_2$ (dichloramine) $+ H_2O$

4) $NH_2Cl + HOCl \rightarrow NCl_3$ (nitrogen/trichloride) $+ H_2O$

5) $NH_4^+ + 4HOCl \rightarrow NO_3^-$ (nitrate) $+ 4HCl + H_2O + H^+$

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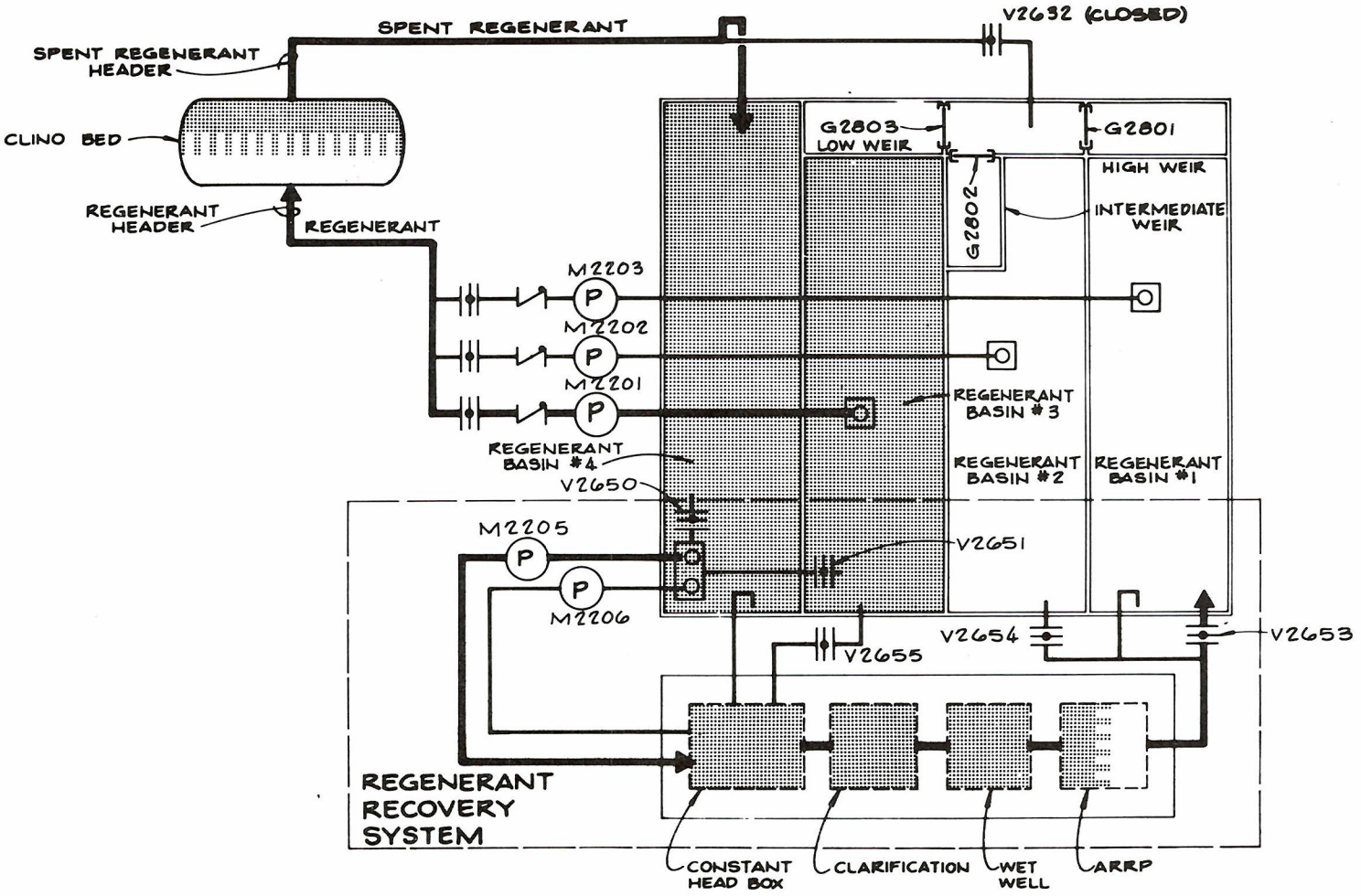


FIG 3
AMMONIA ELUTRIATION CYCLE - PHASE 1

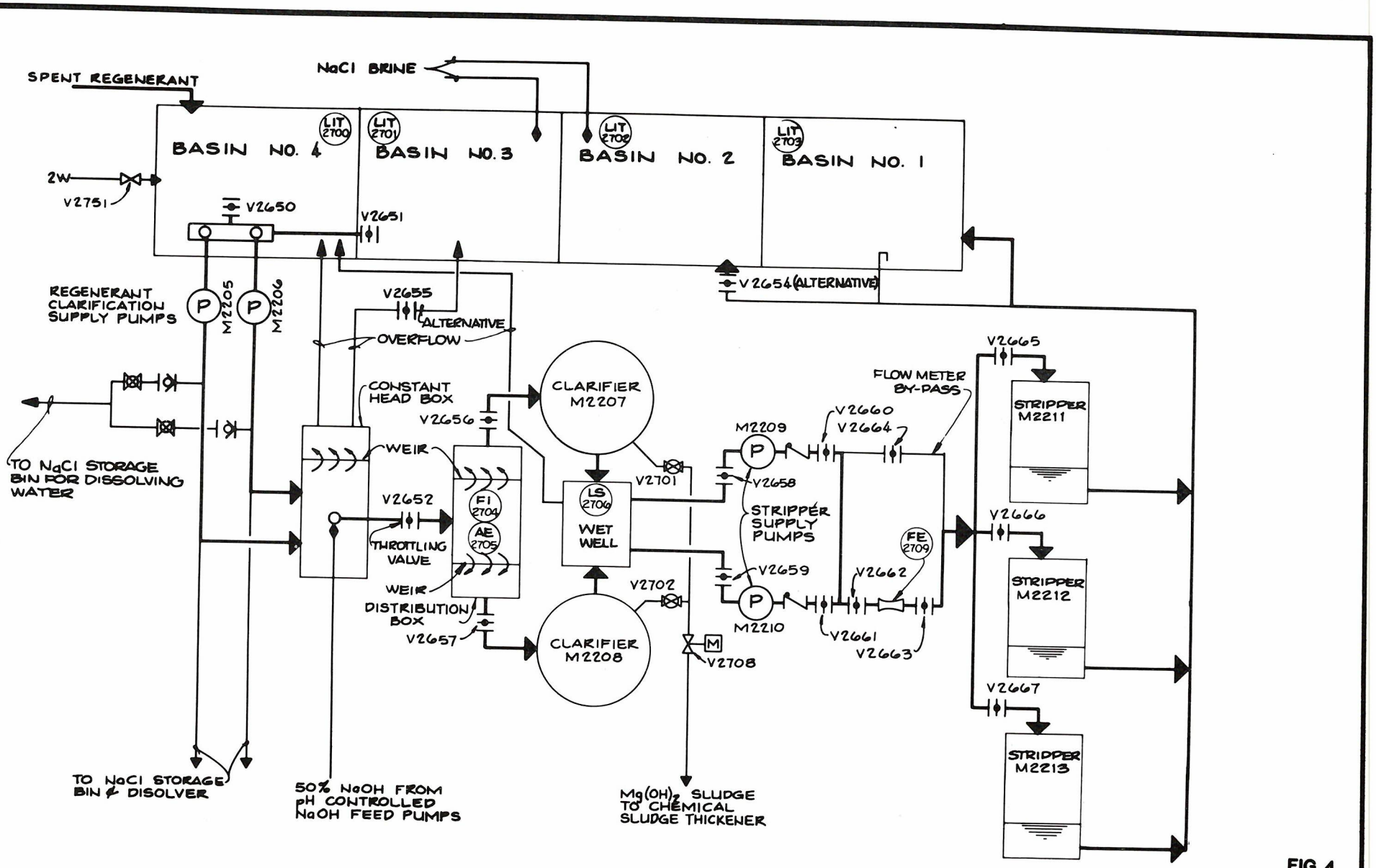


FIG 4
SCHEMATIC OF REGENERANT CLARIFICATION & ARRP STRIPPER PROCESS

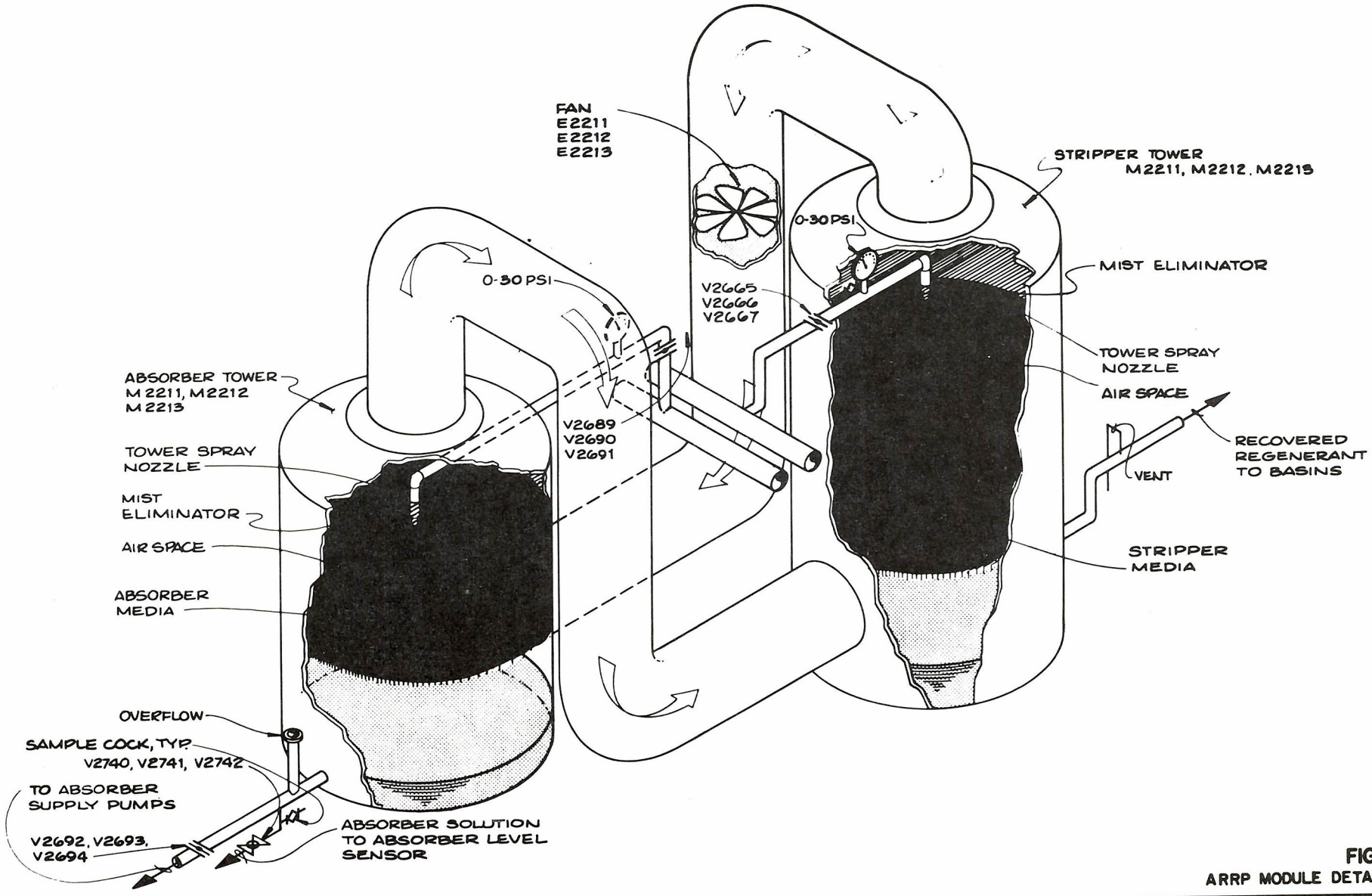
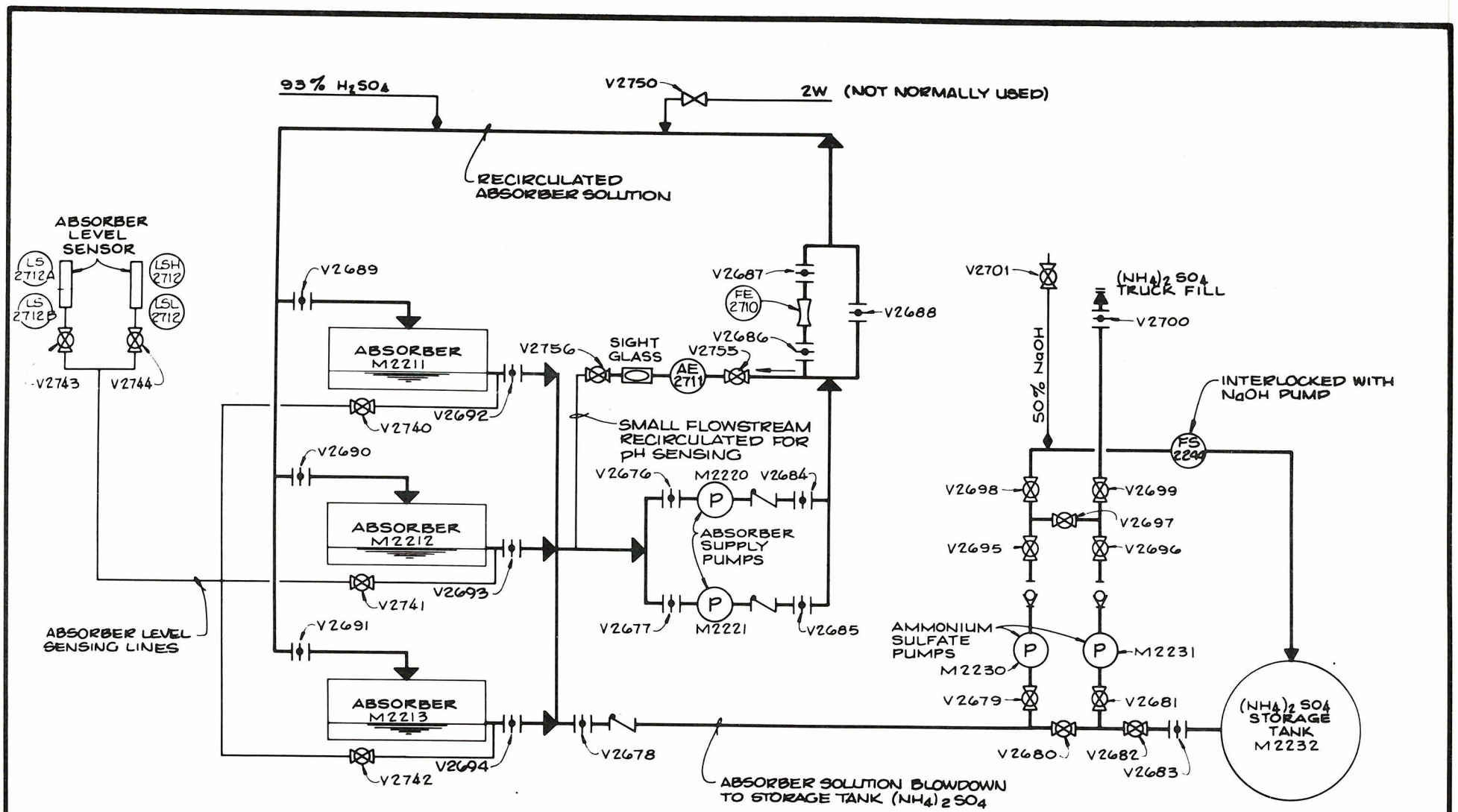
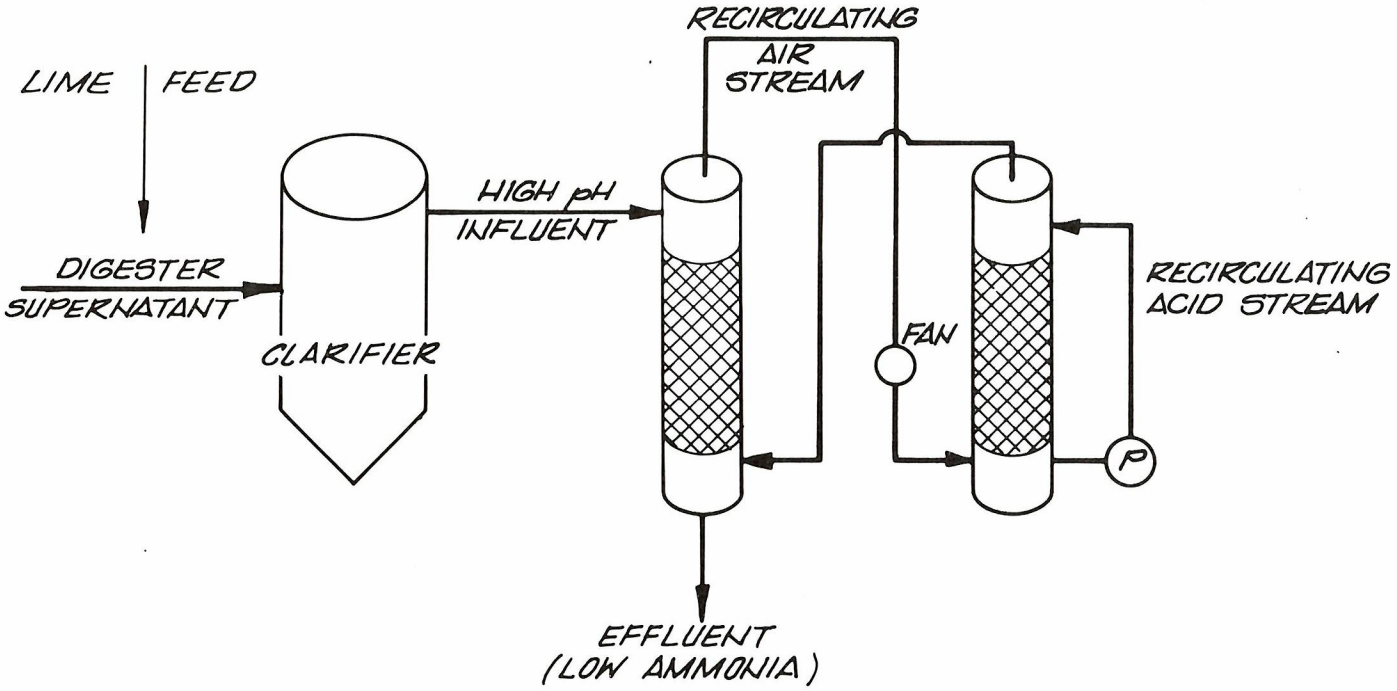


FIG 5
ARRP MODULE DETAILS



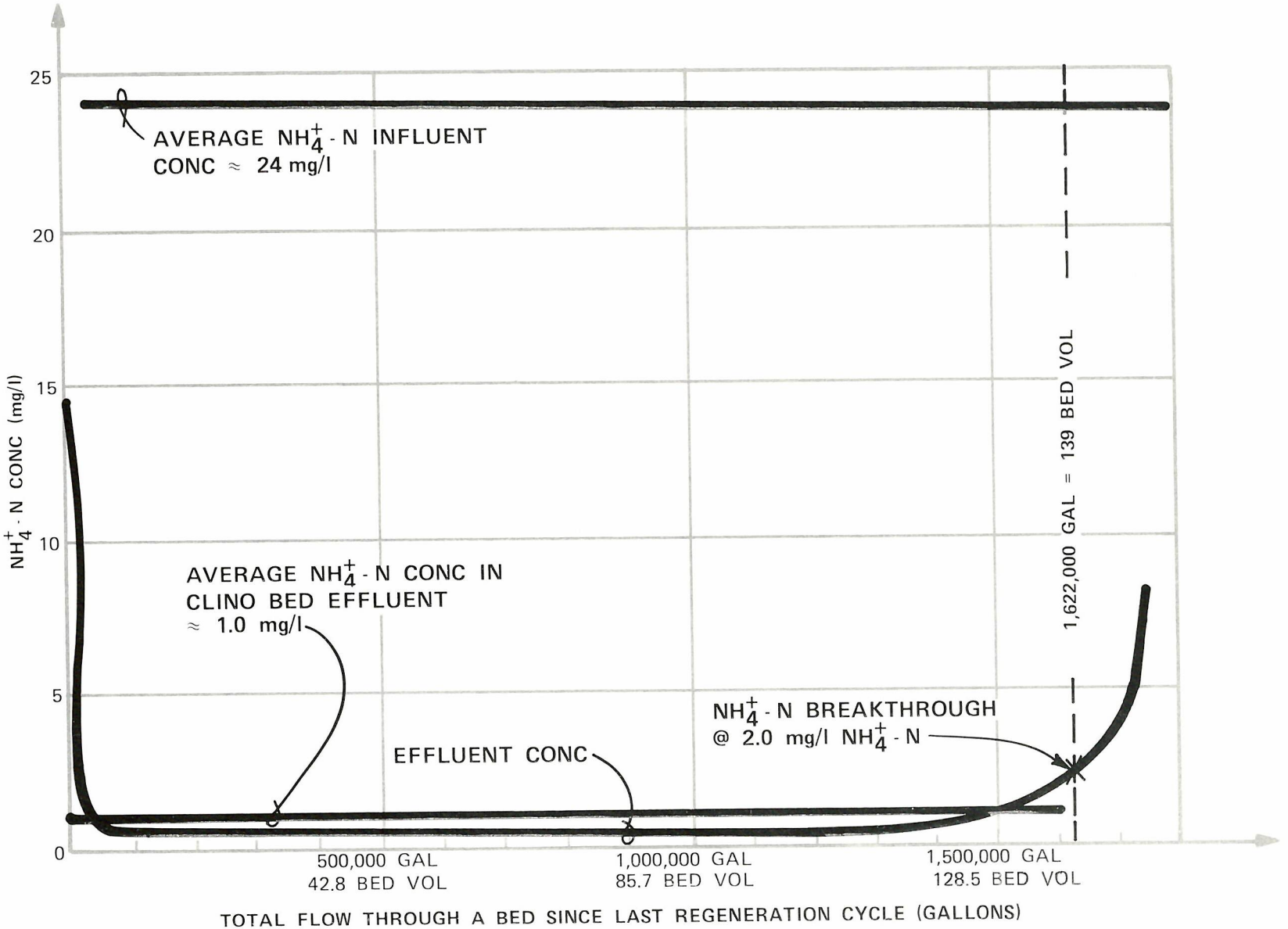
SCHEMATIC OF ARRP ABSORBER SYSTEM & AMMONIUM SULFATE TRANSFER SYSTEM

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ARRP PROCESS FLOW DIAGRAM

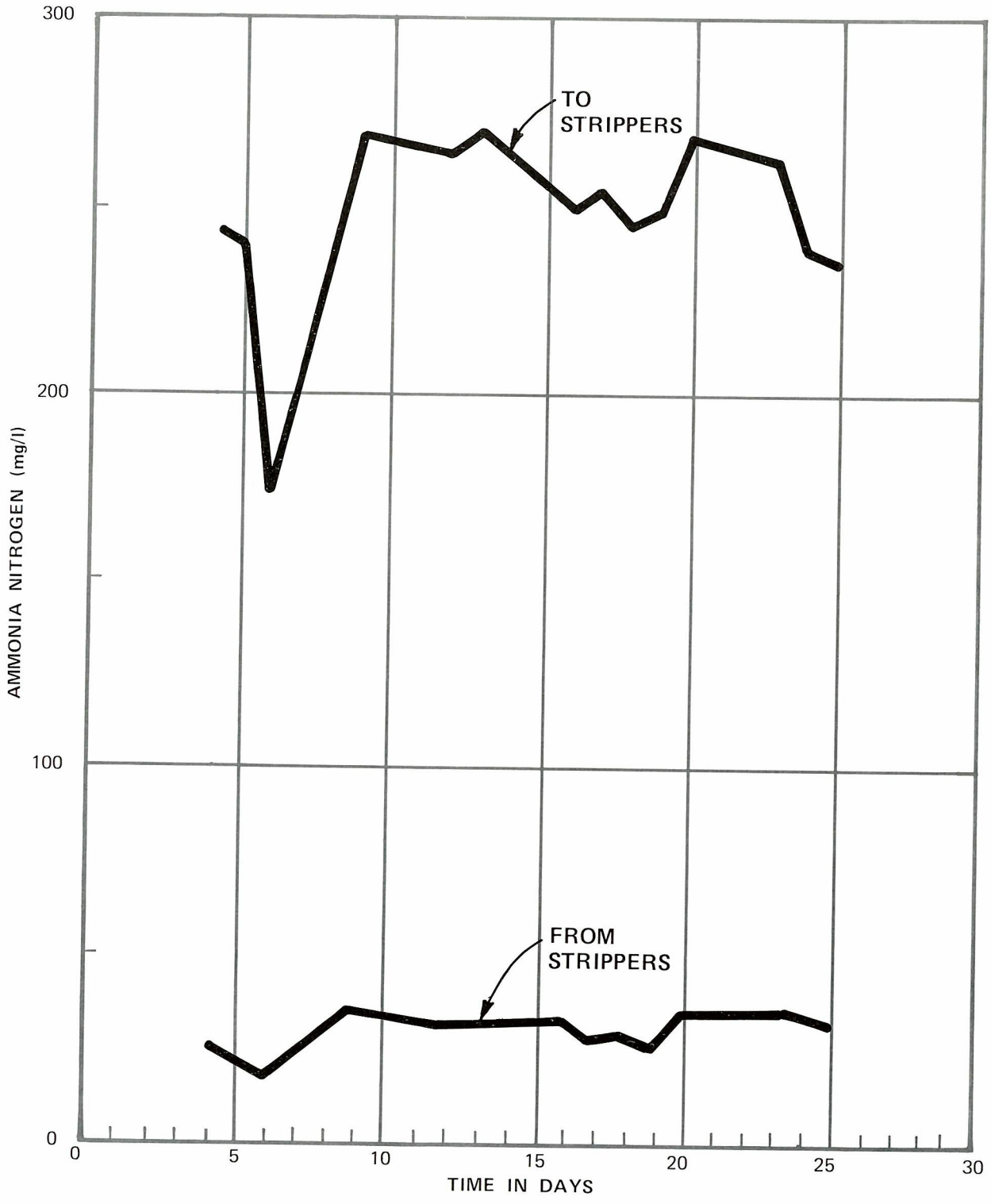
FIGURE 5



■ T-TSA AWT PLANT
CLINO BED EFFLUENT AMMONIA CONCENTRATION AS A FUNCTION OF BED THROUGHPUT

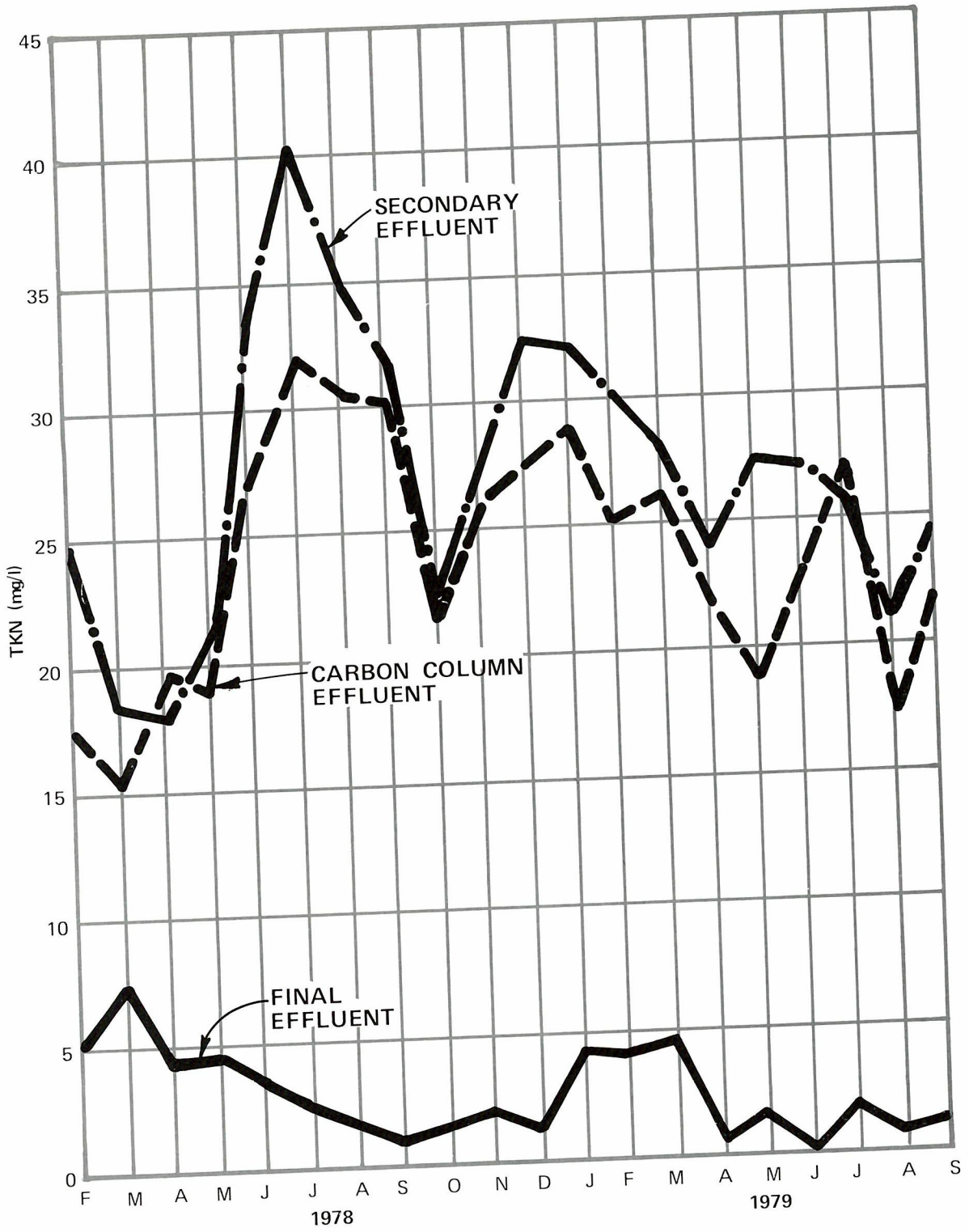
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T-TSA AWT PLANT
AMMONIA REMOVAL FROM SPENT REGENERANT
BY CLOSED CYCLE AIR STRIPPING PROCESS
APRIL 1979

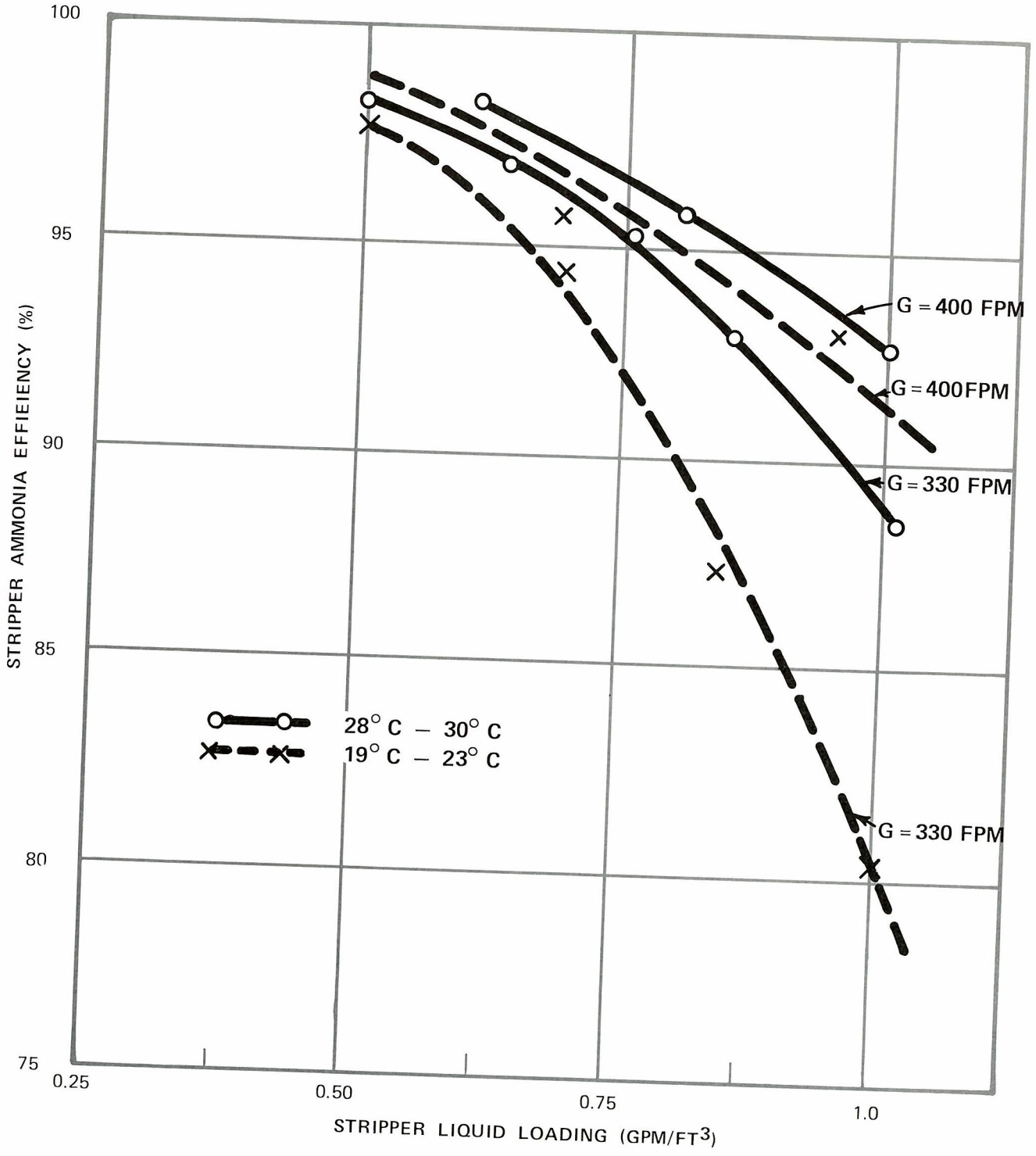
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■ **T-TSA AWT PLANT**
MONTHLY AVERAGE TKN
1978-1979

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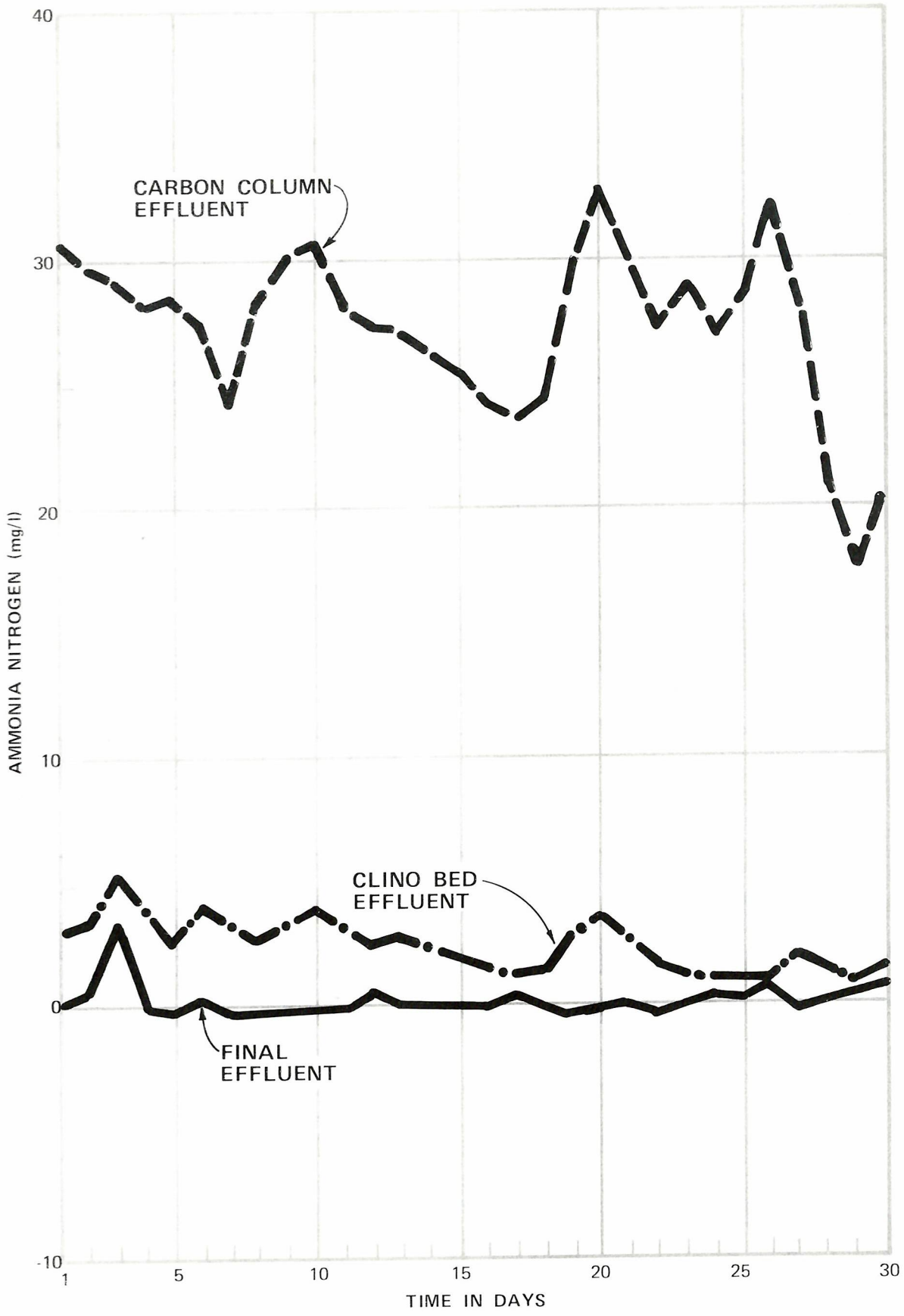
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T-TSA AWT PLANT
CLOSED CYCLE AIR STRIPPING REGENERANT RENEWAL PROCESS EFFICIENCY AS A FUNCTION OF STRIPPER HYDRAULIC LOAD AND AIR FLOW

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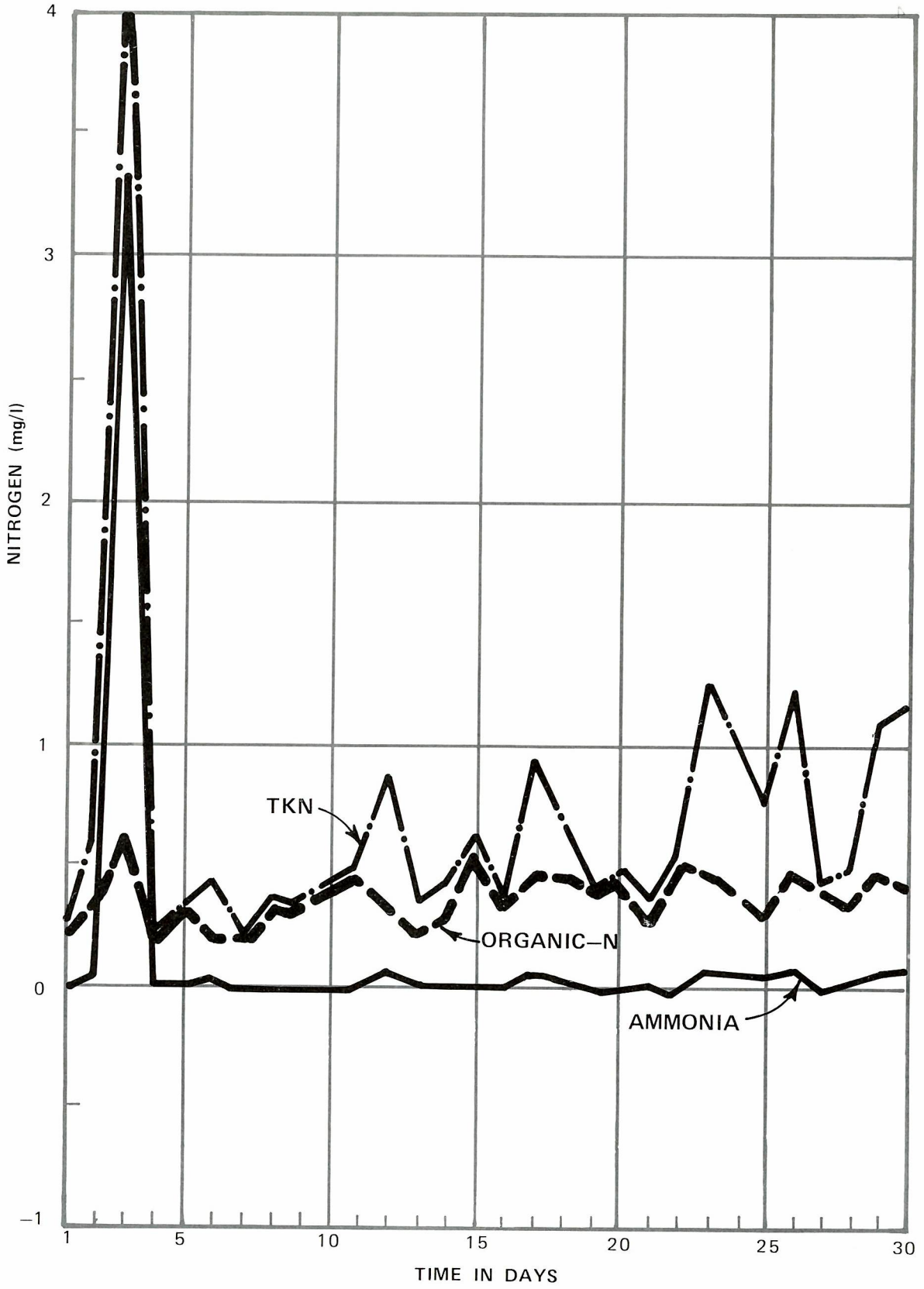
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T-TSA AWT PLANT
EFFLUENT NH_3-N BY UNIT PROCESS
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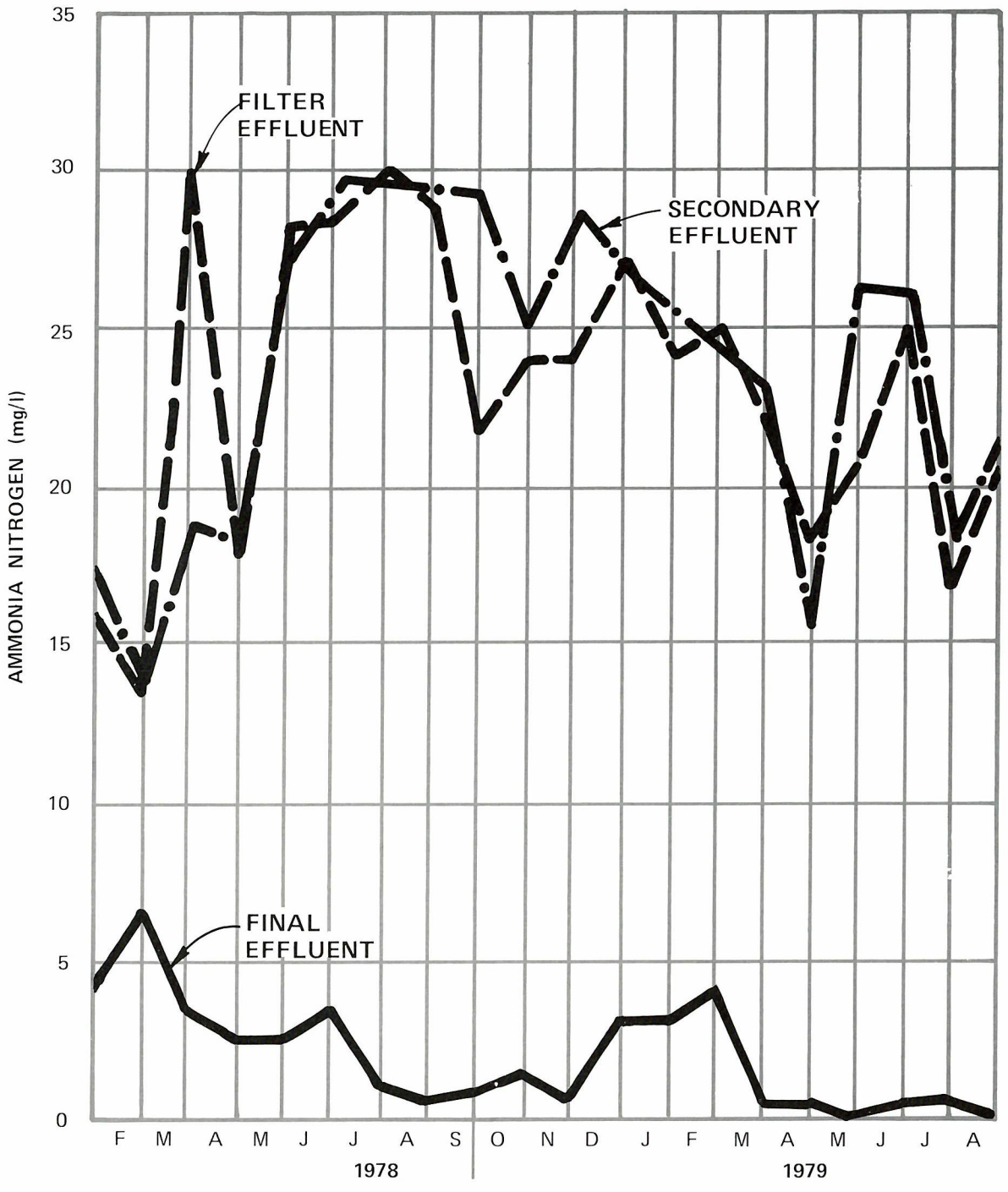
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T-TSA AWT PLANT
FINAL EFFLUENT NITROGEN
APRIL 1979

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T-TSA AWT PLANT
MONTHLY AVERAGE NH_3-N
1978-1979

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NITRIFICATION SYSTEMS WITH INTEGRATED PHOSPHORUS PRECIPITATION

by

P.M. Sutton, K.L. Murphy and B.E. Jank

Presented at the

PCAO Conference
Toronto, Ontario
April 18, 1977

and

B.C. Water and Waste Association Conference
Harrison Hot Springs, B.C.
April 19, 1977

INTRODUCTION

The desire to produce high quality effluents from municipal wastewater treatment plants has led to increasingly stringent limitations on the discharge of phosphorus and ammonia-nitrogen. The addition of iron or aluminum salts to a nitrifying activated sludge plant represents an economical method for phosphorus and nitrogen control.

Several activated sludge process alternatives are available for nitrogen transformation and may be classified as either combined or separate sludge carbon removal-nitrification systems. Recent pilot plant studies under constant flow conditions have indicated that equal degrees of nitrification can be achieved in the combined and separate sludge systems provided the sludge age or solids retention time (SRT) of the systems is maintained at a level to ensure an adequate population of nitrifiers (Sutton, Murphy, Jank and Monaghan, 1975, Lawrence and Brown, 1976). In the design of a system for both nitrogen and phosphorus control, an important consideration in the selection of a combined or separate sludge system is the compatibility of the chemical precipitant and the required microbial population. Although several investigators have studied phosphorus removal by precipitation with metal salts, little attention has been paid to the effects on nitrification.

The optimum pH condition for phosphorus removal by aluminum and iron is normally reported around 5 to 6. Since the optimum range for carbonaceous oxidation is approximately pH 6 to 8, the two treatment goals of carbon and phosphorus removal can be met simultaneously. As the optimum pH condition for nitrification is around 8 to 9 (Wild, Sawyer and McMahon, 1971), and carbon removal-nitrification and phosphorus removal are to be accomplished simultaneously, a compromise

in establishing an operating pH must be reached.

Nitrification is a hydrogen ion producing reaction. Assuming the pH is less than 8.5, the hydrogen ions produced react with the bicarbonate in the wastewater, resulting in an increase in CO_2 concentration and a decrease in bicarbonate alkalinity (Table 1). The additional alkalinity demand imposed by the addition of alum or ferric chloride (Table 1) may result in a significant pH reduction unless there is adequate buffering capacity in the wastewater.

The results from the operation of a 132 m^3/day (20 Igpm)* extended aeration pilot plant with alum or ferric chloride indicated that nitrification could still be maintained (Sutton, Murphy, Jank and Monaghan, 1975) even when the wastewater was moderate in alkalinity (150 mg/l as $CaCO_3$) and the resulting reactor pH was less than 7. Similarly Hais, Stamberg and Bishop (1971), reported that substantial nitrification occurred in a step feed activated sludge system with alum addition even though the pH in the final stage was less than 7.

Black (1971) reported successful operation of a carbon removal-nitrification extended aeration system when alum or ferric chloride was added for phosphorus removal. The experimental program was not designed to study the effects of chemical addition on nitrification and therefore the results were based on a very limited amount of data.

Contrary to these reports, Long, Nesbitt and Kountz (1971), operating a portion of the Pennsylvania State University biological wastewater treatment plant, reported a decrease in nitrification when alum was used as the precipitating chemical. The lower degree of nitrification corresponded to a reduction in pH which occurred as a result of addition of alum into the effluent end of the aeration tank

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* Imperial gallons per minute

TABLE 1. EFFECTS OF NITRIFICATION AND CHEMICAL ADDITION ON REACTOR ALKALINITY

BIOLOGICAL OXIDATION AND NEUTRALIZATION

$$NH_4^+ + 1.5\ O_2 \rightarrow 2\ H^+ + H_2O + NO_2^- \quad \ldots(1)$$

$$H^+ + HCO_3^- \rightarrow CO_2 + H_2O \quad \ldots(2)$$

PHOSPHORUS PRECIPITATION AND HYDROLYSIS

Alum

$$Al_2(SO_4)_3 \cdot 14H_2O + 2H_2PO_4^- + 4HCO_3^- \rightarrow 2AlPO_4\downarrow + 3SO_4^{-2} + 14H_2O + 4H_2CO_3 \quad \ldots(3)$$

$$Al_2(SO_4)_3 \cdot 14H_2O + 6HCO_3^- \rightarrow 2Al(OH)_3\downarrow + 3SO_4^{-2} + 14H_2O + 6CO_2 \quad \ldots(4)$$

Ferric Chloride

$$FeCl_3 \cdot 6H_2O + H_2PO_4^- + 2HCO_3^- \rightarrow FePO_4\downarrow + 3Cl^- + 6H_2O + 2H_2CO_3 \quad \ldots(5)$$

$$FeCl_3 \cdot 6H_2O + 3HCO_3^- \rightarrow Fe(OH)_3\downarrow + 3Cl^- + 3CO_2 + 6H_2O \quad \ldots(6)$$

When alum was added at the head end of the tank no reduction in pH or nitrification was reported but since the solids retention time (SRT) during this period was much higher, the authors were not able to draw any definite conclusions.

Unz and Davis (1975) found municipal wastewater sludges treated with alum contained fewer nitrite-oxidizing bacteria than untreated sludges. This result could not be conclusively related to aluminum ion toxicity. The sludges had undergone a sharp reduction in pH and had SRT's of less than two days, conditions not conducive to the build-up of nitrifiers. No significant difference was observed between ammonia-oxidizing populations in alum treated and untreated sludges.

In view of the conflicting reports it was considered that additional kinetic and operational data were required to define specifically the effects of chemical addition on nitrifying systems.

EQUIPMENT AND PROCEDURES

Pilot Plant Design and Operation

The carbon removal-nitrification systems employed to evaluate concurrent phosphorus and nitrogen control included a two-stage separate (TSS) sludge system and a two-stage combined (TSC) sludge system (Figure 1). The reactors, designated as plants A and B, consisted of two, 2.18 m^3 (480 Igal) dispersed air aeration tanks. Plant B incorporated an intermediate clarifier separating the reactor into two aeration tanks (B1 and B2). With this intermediate clarifier recycling sludge to the first aeration tank (B1), plant B functioned as a two-stage separate sludge system. With the intermediate clarifier bypassed and sludge recycled from the final clarifier to the first aeration tank, plant B operated as a combined sludge system (TSC) identical to plant A (Figure 1).

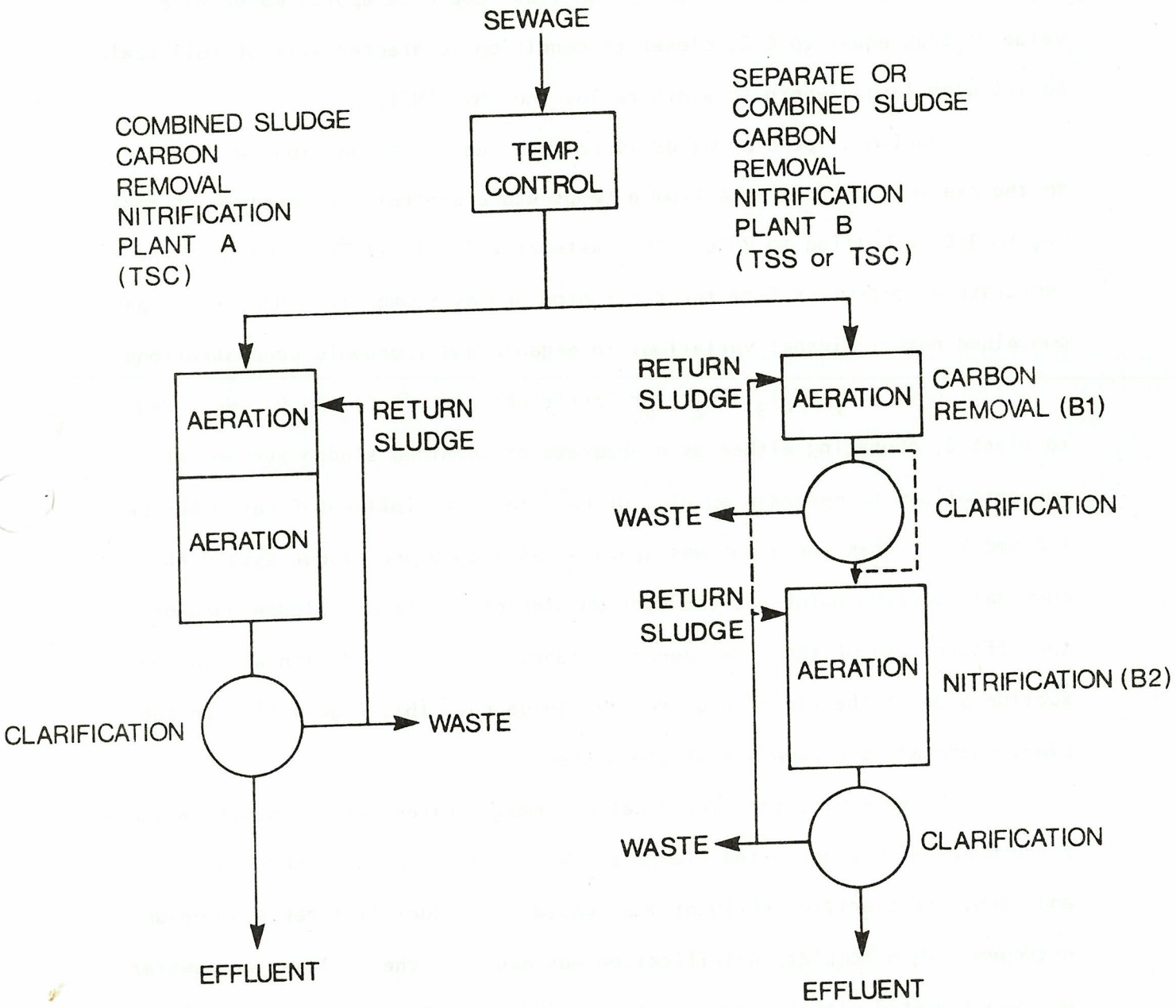


FIGURE 1. FLOW DIAGRAM OF CARBON REMOVAL-NITRIFICATION PILOT PLANT SYSTEMS

The use of dye studies to approximate the hydraulics of the systems indicated that the two aeration tanks of plant B, when operated as a separate sludge system (TSS), could be considered completely mixed, characterized by a large value of the dimensionless dispersion parameter D/uL. The mixing regime in the two-stage combined sludge system (TSC) could be approximated by a value of D/uL equal to 0.2, closer to conditions characteristic of full-scale basins with large length to width ratios (Murphy, 1971).

Variable speed pumps delivered the degritted municipal wastewater to the reactors at 1.25 lgpm from a temperature control unit capable of cooling to 3°C or heating to 25°C. The wastewater (Table 2) was pumped to the temperature controller from the Burlington Skyway Sewage Treatment Plant and contained normal diurnal variations in organic and inorganic concentrations.

Alum ($Al_2(SO_4)_3 \cdot 14H_2O$) or ferric chloride ($FeCl_3 \cdot 6H_2O$) was added to plant B, operating either as a separate or combined sludge system, at concentrations to maintain an Al^{+3} or Fe^{+3} to total influent P ratio between 1.0 and 2.0. When the plant was operated as a combined sludge system the chemical addition point was the suction side of the return sludge pump or at the effluent end of the final aeration tank. Chemical addition was to the suction side of the first stage return sludge pump (B1, Figure 1) when the plant operated as a separate sludge system.

To determine the nitrification removal rates used to describe the performance of the activated sludge systems with and without chemical addition, the clarified effluent must contain residual filterable ammonium nitrogen. When complete nitrification was expected the influent wastewater was supplemented with inorganic nitrogen (NH_4Cl). Bicarbonate ($NaHCO_3$) was added to the feed when necessary to avoid any alkalinity limitation on the nitrification rate. The dissolved oxygen in the reactors was maintained

TABLE 2. RAW WASTEWATER CHARACTERISTICS

Characteristic mg/l	Mean	90%*
COD	220	315
BOD_5	98	138
SS	135	300
TOTAL P	5	10
NH_4-N	12	19
FILTERABLE TKN	15	21
ALKALINITY (as $CaCO_3$)	231	269

*90% of observations are equal to or less than stated value.

at or above 2.0 mg/l.

The performance of the nitrification systems was assessed using 24 hr. flow-weighted sampling under temperature and system solids retention time conditions specified by the experimental design. The SRT, defined as the suspended solids under aeration divided by the daily solids lost in the effluent and through sludge wasting, was controlled automatically by wasting from the return sludge lines. Calculation of the SRT for the separate sludge system involves consideration of the individual SRT's for the two aeration tanks (B1 and B2, Figure 1). Details of the calculation are given elsewhere (Sutton, Murphy, Jank, and Monaghan, 1977).

Experimental Design

The two-stage combined and separate sludge systems (TSC and TSS) were operated according to the experimental design with two different chemical precipitants and two levels of temperature in order to study the effects of these variables on nitrification. The chemicals selected, alum and ferric chloride, are those most commonly used for integrated phosphorus removal in a biological system. The temperature levels 5° and 25°C reflect the range over which treatment plants may operate in Canada. A parallel two-stage combined sludge system without chemical addition (TSC*) was operated under identical temperature and biological SRT conditions throughout the experimental period.

In operating biological systems, especially with chemical addition, it is most important that equilibrium conditions are attained prior to an investigative period. In all experiments the fraction of the theoretical chemical equilibrium achieved (Figure 2) generally was in excess of 90%, the operating period prior to testing exceeding two times the system solids retention time.

The biological equilibrium in both parallel systems (Plants A and B),

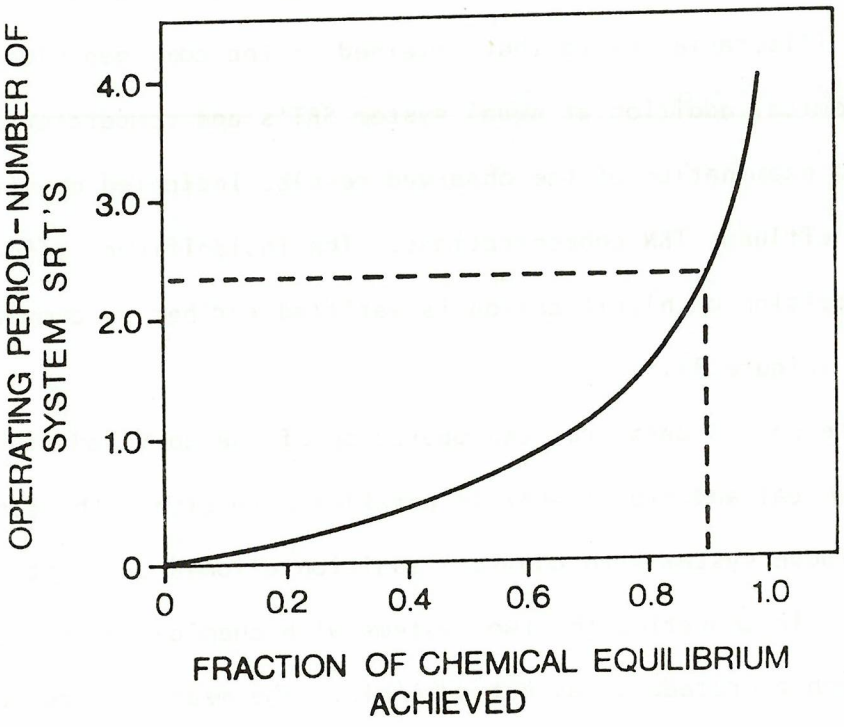


FIGURE 2. THEORETICAL RELATIONSHIP BETWEEN CHEMICAL EQUILIBRIUM AND PERIOD OF OPERATION

was always in excess of the chemical equilibrium as the system SRT's were maintained between 8 and 10 days throughout the experimental program.

DISCUSSION OF RESULTS

Effects of Chemical Addition on Nitrification

The parallel pilot plant configuration (Figure 1) allowed collection of paired data from operation with and without chemical addition for phosphorus removal.

The carbon removal-nitrification systems were compared first in terms of the effect of ferric chloride or alum on the removal of filterable TKN. The paired data indicated comparable removal for both TSC and TSS systems of filterable TKN to that obtained in the combined sludge system without chemical addition at equal system SRT's and temperatures (Figure 3). Statistical examination of the observed results indicated no difference in filterable effluent TKN concentrations. The insignificant effect of chemical addition on nitrification is verified further by comparable nitrate production (Figure 3).

The paired data from the operation of the combined sludge system without chemical addition (TSC*) in parallel with either the combined or separate sludge system with chemical addition allowed a direct statistical assessment. In comparing the two systems with chemical addition (TSC and TSS) such a procedure was not possible. The mean differences in effluent filterable TKN between both the combined sludge (TSC) system with chemical addition and the separate sludge (TSS) system with chemical addition, and the combined sludge system without chemical addition (TSC*) were determined. The two mean values were examined statistically (Snedecor and Cochran, 1967) to determine if they were significantly different from each other. The resulting "t" value was not significant, implying that equal filterable TKN

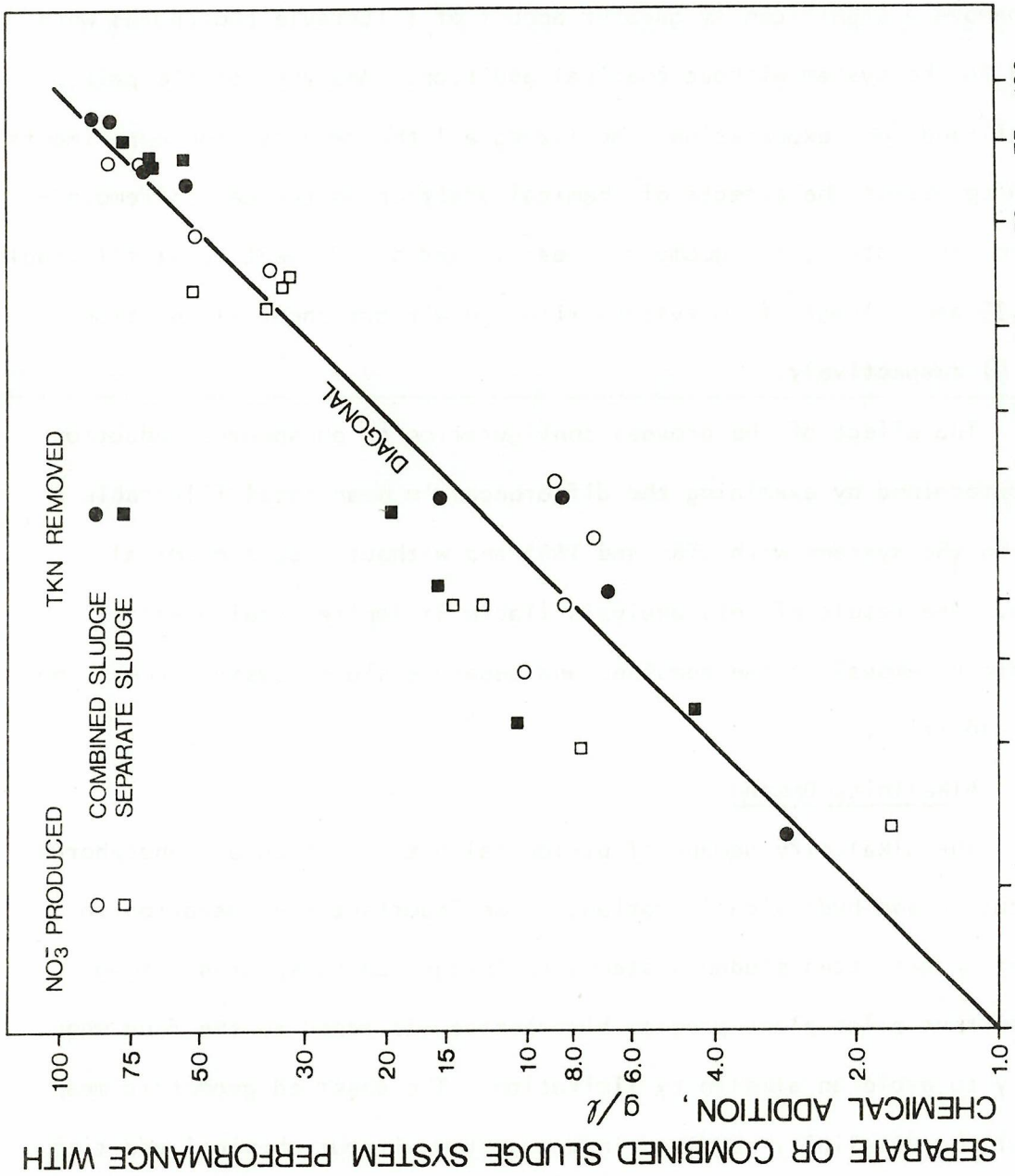


FIGURE 3: PAIRED DATA COMPARISON OF NITRIFICATION PERFORMANCE WITH AND WITHOUT CHEMICAL ADDITION

removal was obtained with both combined (TSC) and separate (TSS) sludge systems receiving chemical addition for phosphorus removal (Table 3). A similar data analysis indicated that both sludge systems (TSC and TSS) produced equal amounts of nitrate (Table 3).

The combined and separate sludge systems receiving chemical addition should remove a significantly greater amount of filterable phosphorus when compared to the system without chemical addition. Analyses of the paired data confirmed this expectation. Analyzing all the results from experiments designed to assess the effects of chemical addition on the carbon removal-nitrification systems, the geometric mean values of effluent total filterable P are 0.35 and 2.1 mg/l from systems with and without chemical addition (Figure 4) respectively.

The effect of the process configuration on phosphorus reduction can be determined by examining the differences in mean total filterable P removed in the systems with (TSC and TSS) and without (TSC*) chemical addition. The result of this analysis (Table 3) implies equivalent filterable P removal in the combined and separate sludge systems receiving chemical addition.

Alkalinity Demand

The alkalinity demand of biological nitrification and phosphorus precipitation and hydroxide formation, is an important consideration in designing an activated sludge system with integrated phosphorus removal. In the current pilot plant program bicarbonate was added to the feed when necessary to avoid an alkalinity limitation. The observed geometric mean value for the alkalinity consumed in the system without chemical addition was 6.0 grams of alkalinity (as $CaCO_3$) per gram of filterable TKN removed and for the system receiving chemical addition the value was 8.0 (Figure 5).

TABLE 3 DIFFERENCE IN EFFLUENT RESULTS FROM COMBINED AND SEPARATE SLUDGE SYSTEMS WITH CHEMICAL ADDITION

Parameter	System Compared to TSC*	$\bar{D}$	t	t $\alpha = .95$
Filterable TKN	TSC	3.29	0.26	2.15
	TSS	2.40		
NO_3^--N	TSC	-0.09	0.80	2.15
	TSS	-2.24		
Total Filterable P	TSC	-1.57	1.01	2.16
	TSS	-1.95		

Note: n = number of paired data points

$$t = \frac{\bar{D}_{TSC} - \bar{D}_{TSS}}{Sp\left(\frac{1}{n_{TSC}} + \frac{1}{n_{TSS}}\right)^{1/2}} \quad \text{where,}$$

S_p^2 = pooled variance for $D_{TSC} - D_{TSS}$, and

$$S_p^2 = \frac{n_{TSC}\,(S^2_{D\ TSC}) + n_{TSS}\,(S^2_{D\ TSS})}{n_{TSC} + n_{TSS}}$$

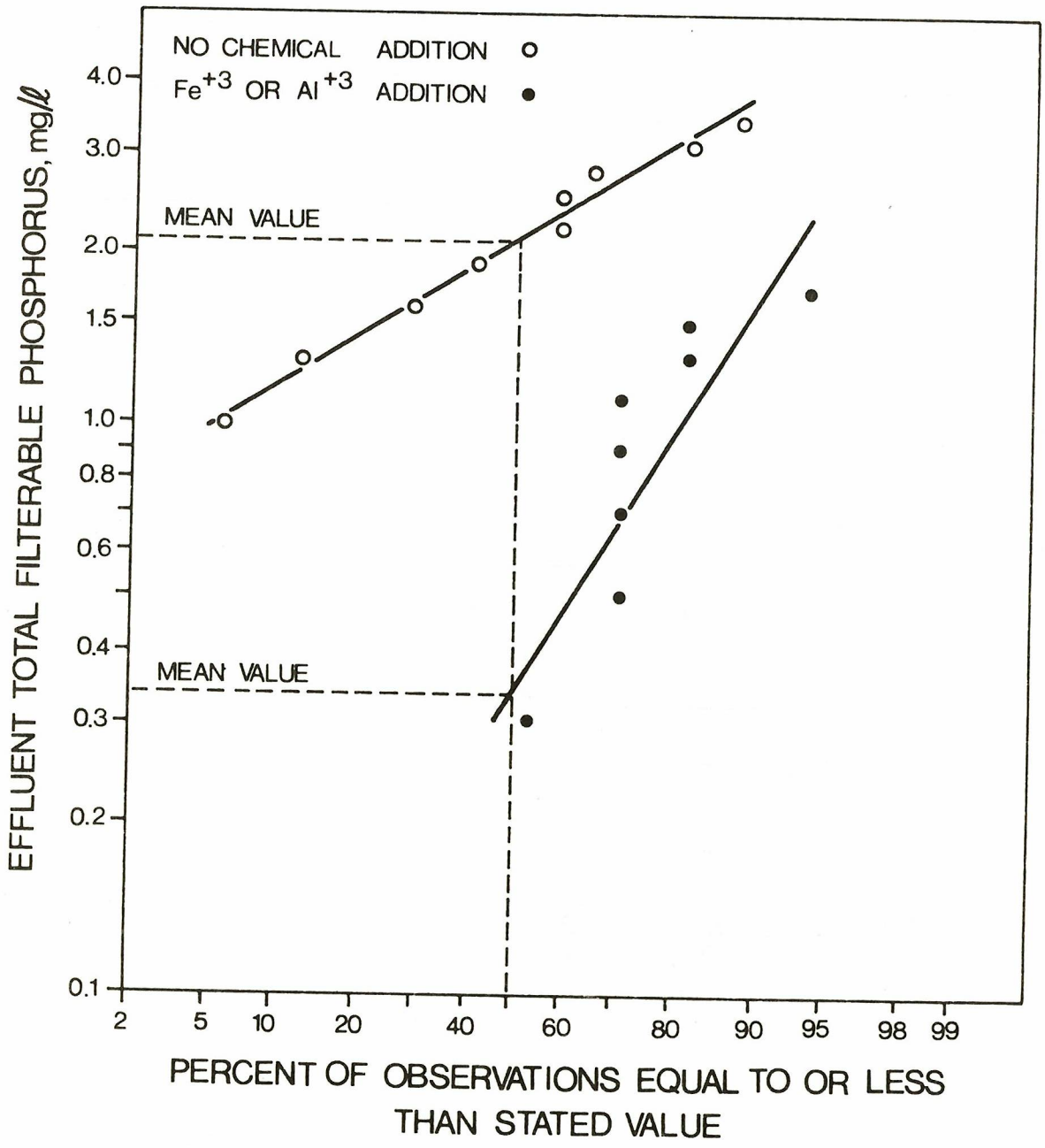


FIGURE 4. PHOSPHORUS REMOVAL IN COMBINED AND SEPARATE SLUDGE SYSTEMS

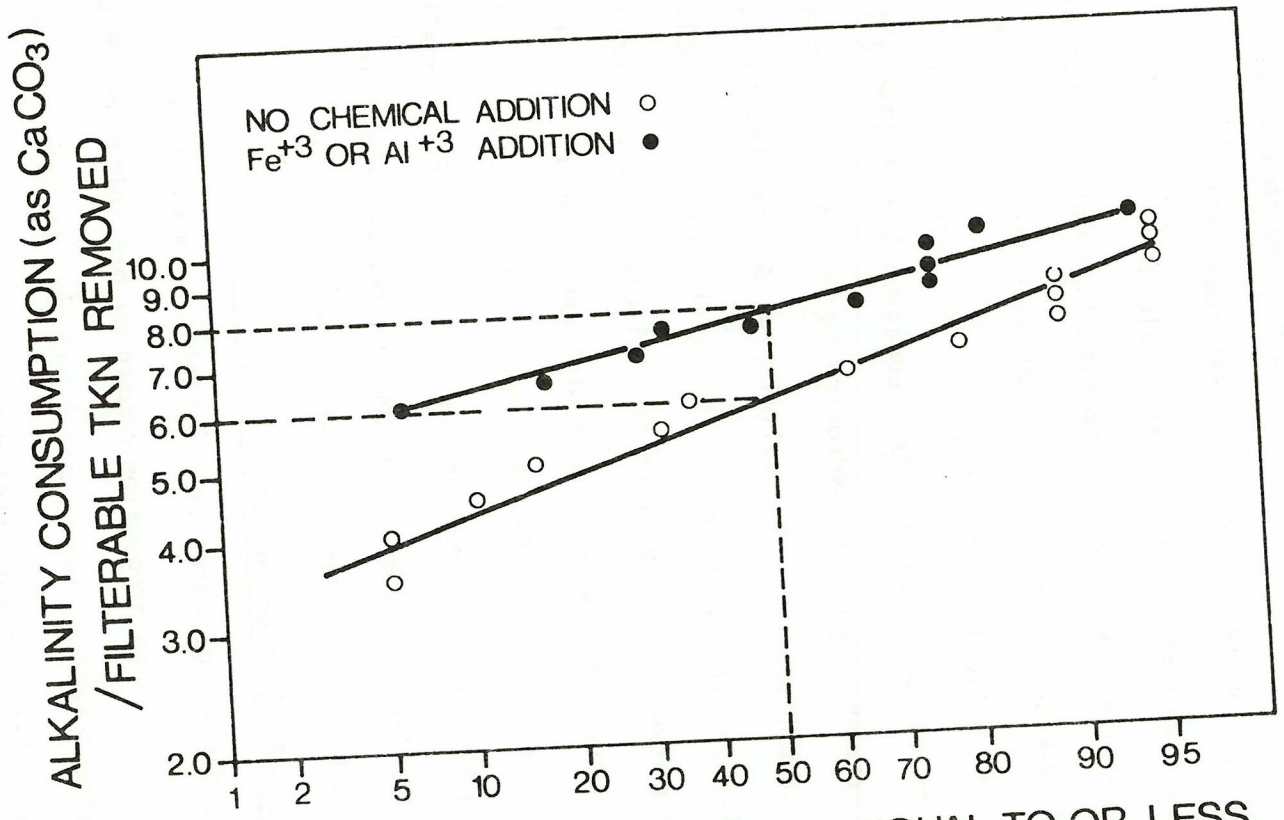


FIGURE 5. ALKALINITY CONSUMPTION IN NITRIFICATION SYSTEMS WITH AND WITHOUT CHEMICAL ADDITION

The stoichiometric alkalinity consumption ratio for nitrification is 7.2 (Table 1) assuming the decrease in filterable TKN is proportional to the decrease in NH_4^+-N.

Parallel operation of the systems with and without chemical addition for phosphorus removal allowed the observed increase in alkalinity consumption to be compared with the calculated theoretical increase. As the paired data analysis indicated that equal filterable TKN removal could be expected from combined and separate sludge systems with and without chemical addition, the observed increase in alkalinity consumption can be directly attributed to the processes of phosphorus precipitation and hydroxide formation. The stoichiometric alkalinity and chemical requirements for precipitation can be calculated from equation 3 or 5 (Table 1). The additional alkalinity consumption due to hydroxide formation can be calculated from equation 4 or 6 (Table 1). The validity of determining the increase in alkalinity consumption by use of the stoichiometric equations was assessed by comparing the observed and calculated values during the parallel operation of combined sludge systems with and without Fe^{+3} addition for a 16 consecutive day period. The mean value for the ratio of observed to theoretical alkalinity consumption was 1.10 (Figure 6). Details of the calculation involved in determining the theoretical alkalinity consumption were presented by Sutton, Murphy, Jank and Monaghan (1977).

Sludge Production and Characteristics

Integrated phosphorus precipitation in a biological system results in an increase in sludge production caused by precipitation (equations 3 and 5) and hydroxide formation (equations 4 and 6). By determining the cumulative solids wasted during a pseudo "steady-state" operating period either in the process effluent or by intentional daily wasting, solids production from the parallel operating alternatives was compared.

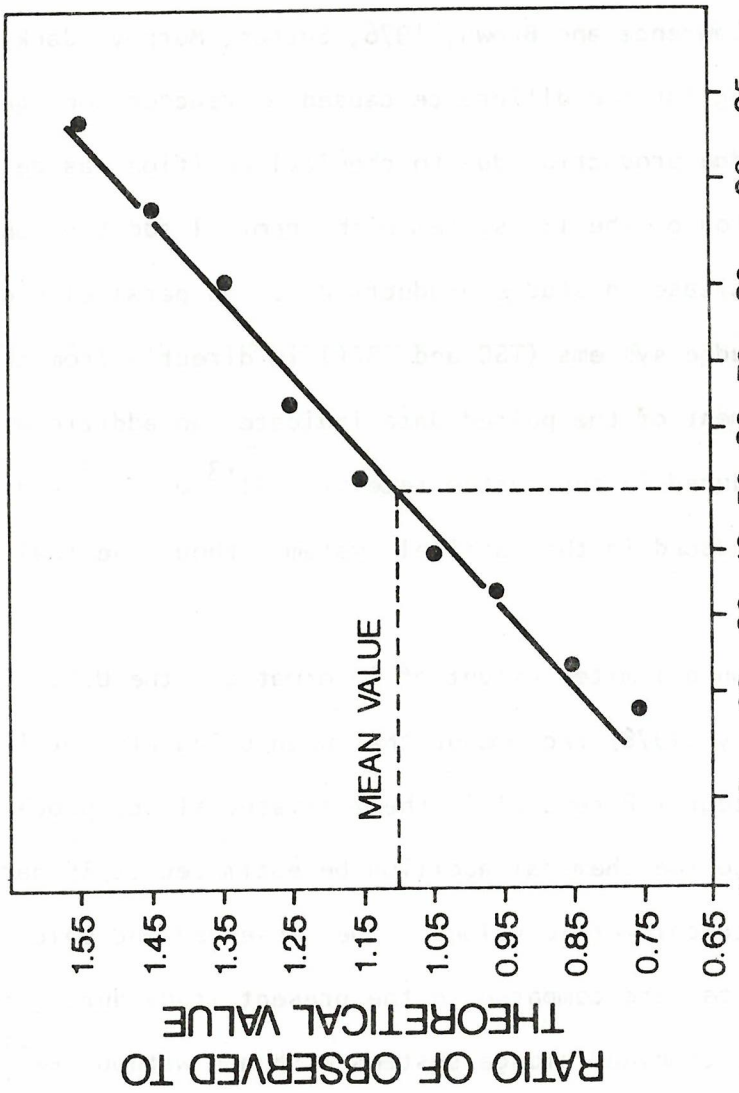


FIGURE 6. ALKALINITY CONSUMPTION DUE TO FERRIC IRON ADDITION

Any increase in sludge production during operation of the separate sludge system (TSS) with chemical addition in parallel with the combined sludge system without chemical addition (TSC*), will be caused by the effects of both chemical addition and reactor configuration. The low solids retention time of the first stage reactor (reactor B1, Figure 1) of the separate sludge system, results in a significantly greater amount of biological solids production in this system in comparison to the combined sludge system (Lawrence and Brown, 1976, Sutton, Murphy, Jank and Monaghan, 1975). Adjusting for the difference caused by reactor configuration, the increase in sludge production due to chemical addition was determined during parallel operation of the TSS system with chemical addition and the TSC* system. Any increase in sludge production during parallel operation of the combined sludge systems (TSC and TSC*) is directly from chemical addition.

Assessment of the paired data indicated an additional 1.23 kg of solids were produced in the system receiving Al^{+3} or Fe^{+3} addition for each kg of solids produced in the parallel system without chemical addition (Figure 7).

Based on a limited amount of information, the U.S. Environmental Protection Agency (1976) recommends that when using alum or ferric chloride addition for integral P removal in the activated sludge process, the sludge production due to the chemical addition be estimated at 35 percent in excess of the stoichiometric values. The observed and calculated sludge production results were compared in the present study during the parallel operation of the combined sludge systems with and without Fe^{+3} addition for a 16-day consecutive period to determine the deviation from the stoichiometric equations (Table 1). The calculated sludge production was determined using a procedure similar to that used to determine the theoretical increase

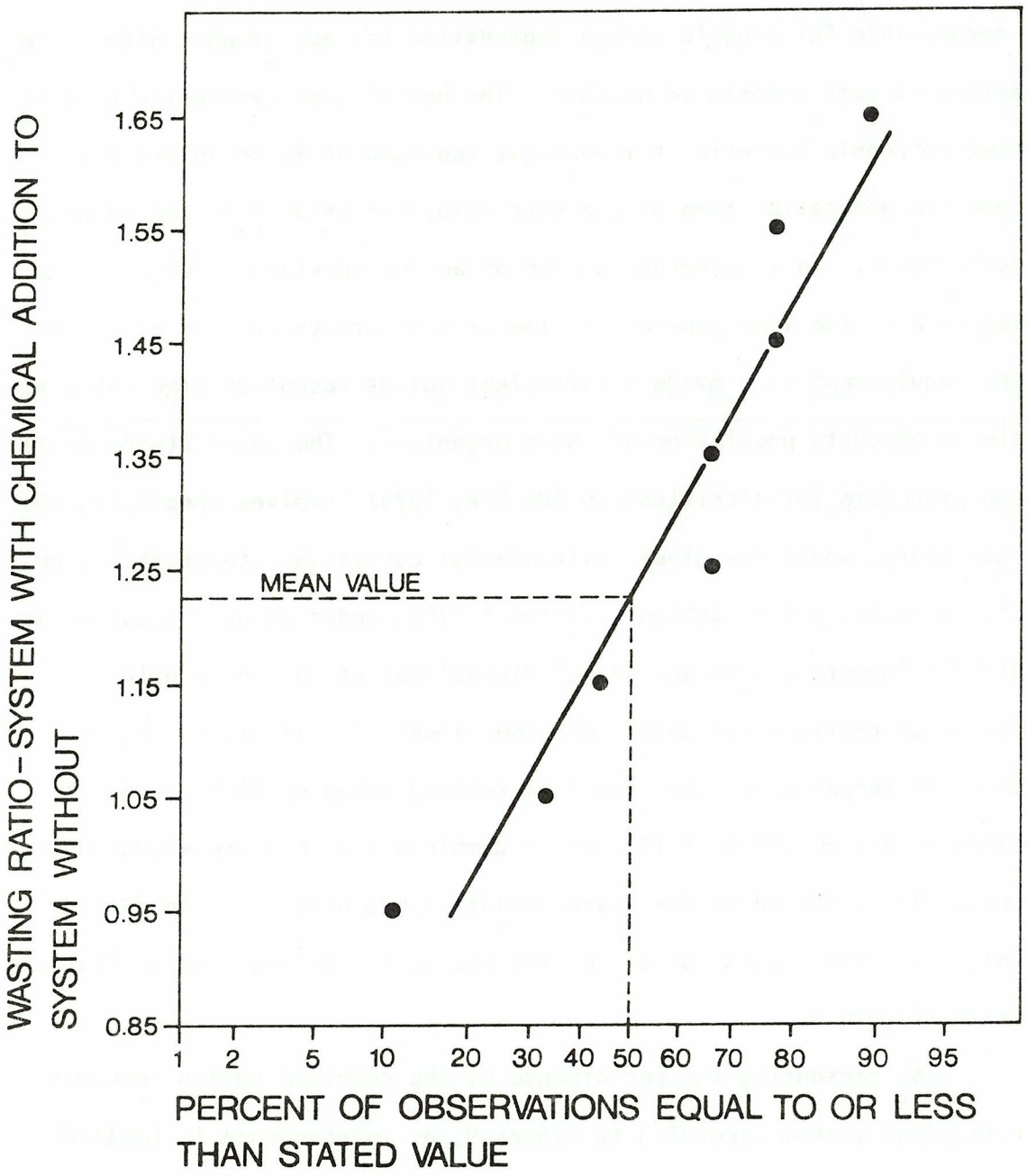


FIGURE 7. INCREASE IN REQUIRED SOLIDS WASTING DUE TO FERRIC IRON OR ALUM ADDITION

in alkalinity consumption due to chemical addition. The mean value of the observed chemical sludge production was 60 percent greater than the stoichiometric value (Figure 8).

Nitrification Rates

In a carbon removal-nitrification system, heterotrophic bacteria are responsible for organic carbon degradation and autotrophic nitrifying organisms convert ammonia to nitrate. The doubling or generation time for the heterotrophic bacteria is frequently reported as 20 to 40 minutes whereas the generation time of the nitrifying bacteria is in the range of 10 to 30 hours. In a suspended growth or activated sludge system a direct consequence of the long generation time or slow growth rate of nitrifiers is the requirement to provide a sufficient solids retention time (SRT) to retain an adequate population of these organisms. The usual steady-state design procedure for nitrification (US EPA, 1975) involves specifying the SRT necessary, under the given environmental conditions (temperature, pH, etc.), to ensure nitrification. Sutton (1976), under pseudo "steady-state" conditions (constant flow and normal diurnal variations in organic and inorganic concentrations) determined the amount of filterable TKN, as a function of temperature, that could be removed based on 24 hr composite sampling at system SRT's of 10 days in combined and separate sludge nitrifying systems. The combined sludge system relationship describes the data derived in this study from operation of the TSC sludge system with and without chemical addition (Figure 9).

By presenting the performance of the combined carbon removal-nitrification system according to Figure 9, an independence is implied between TKN removal and the BOD loading. A significantly different value for the influent BOD from that encountered in this study will result in a

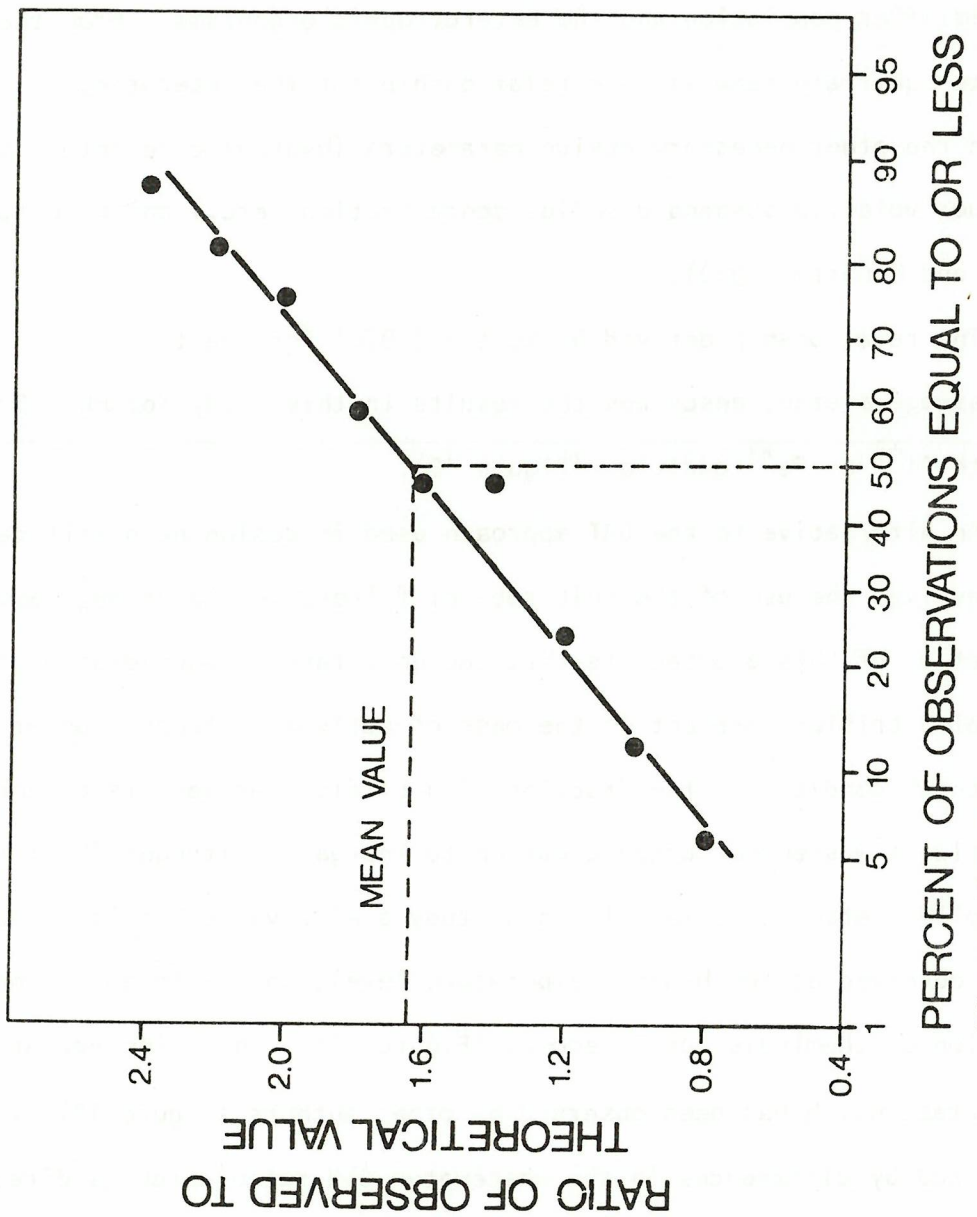


FIGURE 8. SOLIDS PRODUCTION DUE TO FERRIC IRON ADDITION

difference in cell yield (heterotrophs) at the same SRT, and subsequently a difference in the amount of N removed by synthesis. Accounting for this difference, the pseudo "steady-state" design of a nitrifying activated sludge system can be specified by noting the SRT necessary for a given filterable TKN removal. Fixing the SRT establishes the growth rate of both the autotrophic nitrifier population and the heterotrophic organisms. From the growth rate-substrate removal rate relationship for the heterotrophic population the other necessary design parameters (hydraulic retention time, mixed-liquor volatile suspended solids concentration, etc.) can be established (Lawrence and McCarty, 1970).

The relationship derived by Sutton (1976) for the two-stage separate sludge system, describes the results in this study for the TSS system with Al^{+3} or Fe^{+3} addition (Figure 10).

An alternative to the SRT approach used in designing nitrification systems involves the use of the unit rate of filterable TKN or NH_4^+ removal. The limitation of this approach is that the unit rate is dependent on the fraction of nitrifiers present in the mass of activated sludge. Under pseudo "steady-state" conditions, the fraction of nitrifiers present is dependent on the influent wastewater organic carbon to inorganic nitrogen (C:N) ratio and the solids retention time. In this study a wide variation in the unit rates was observed at the higher temperature level, both with and without the addition of chemicals for P removal (Figure 11). This wide variation in unit rates, which has been observed by other authors (Figure 12), is likely caused by differences in the wastewater C:N ratio which is directly reflected by the mixed-liquor suspended solids level required to maintain the same system solids retention time during a pseudo "steady-state" operating period (Figure 13). The higher yield of the heterotrophic

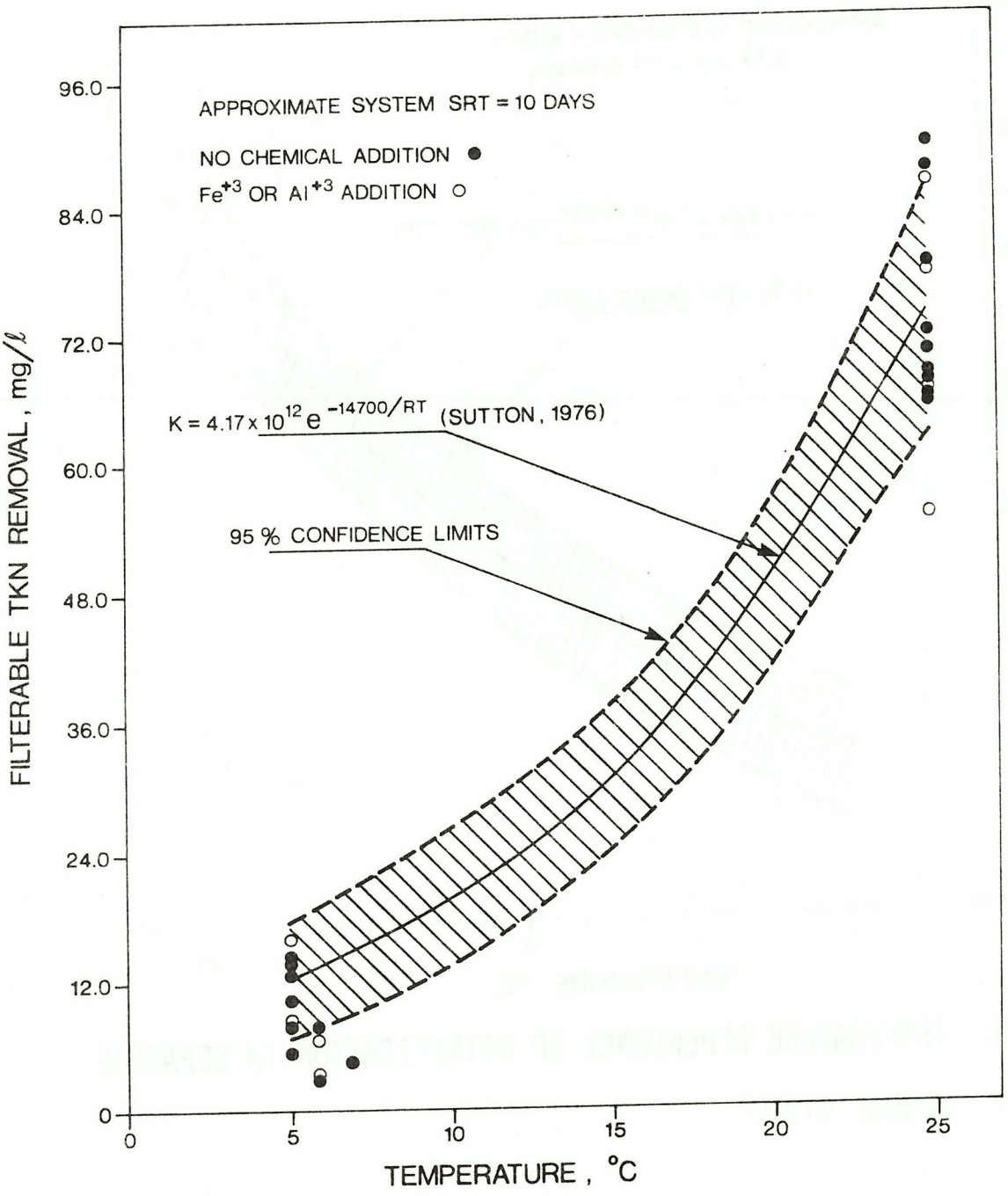


FIGURE 9. TEMPERATURE DEPENDENCY OF NITRIFICATION IN COMBINED SLUDGE SYSTEMS

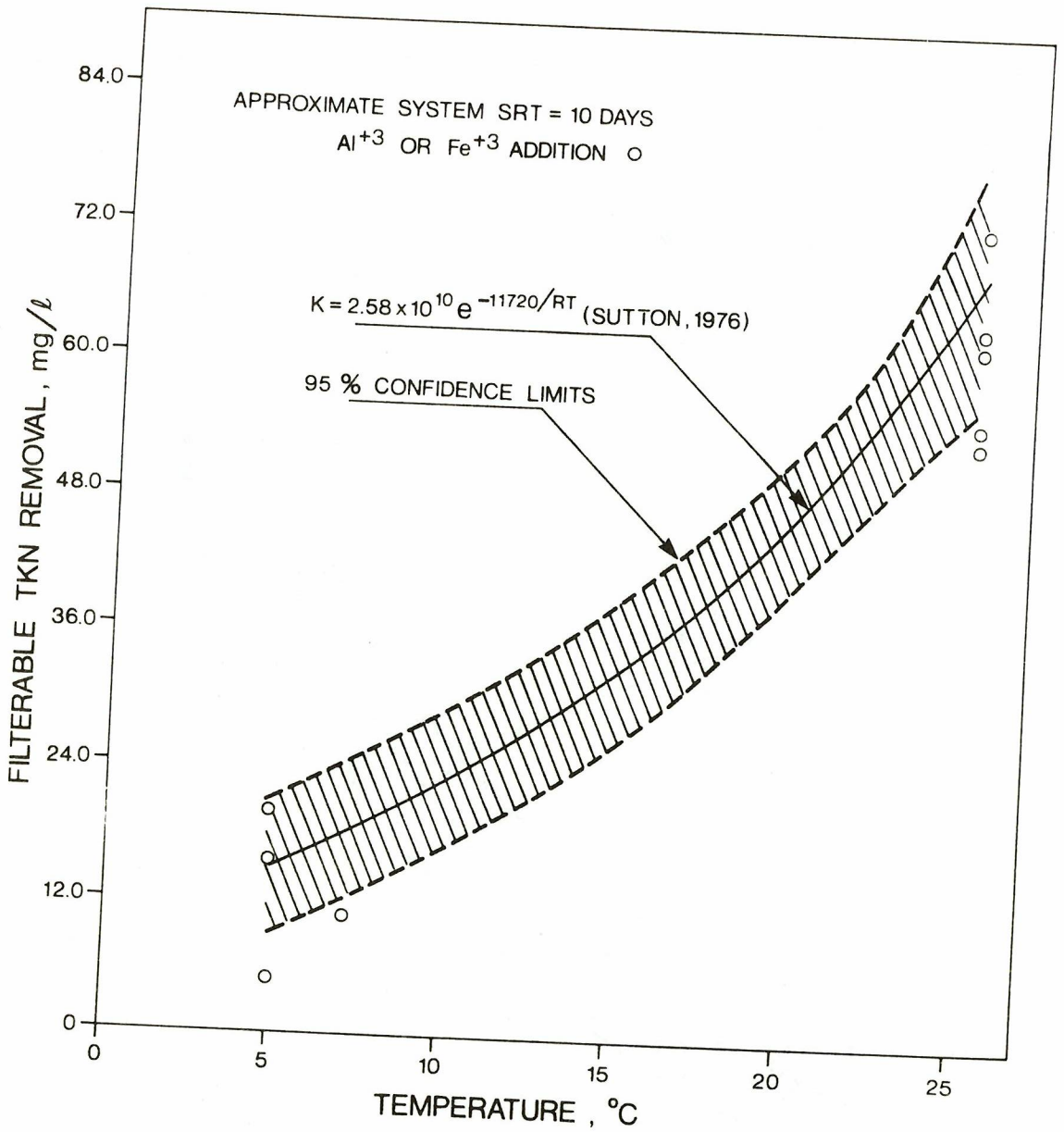


FIGURE 10. TEMPERATURE DEPENDENCY OF NITRIFICATION IN SEPARATE SLUDGE SYSTEM

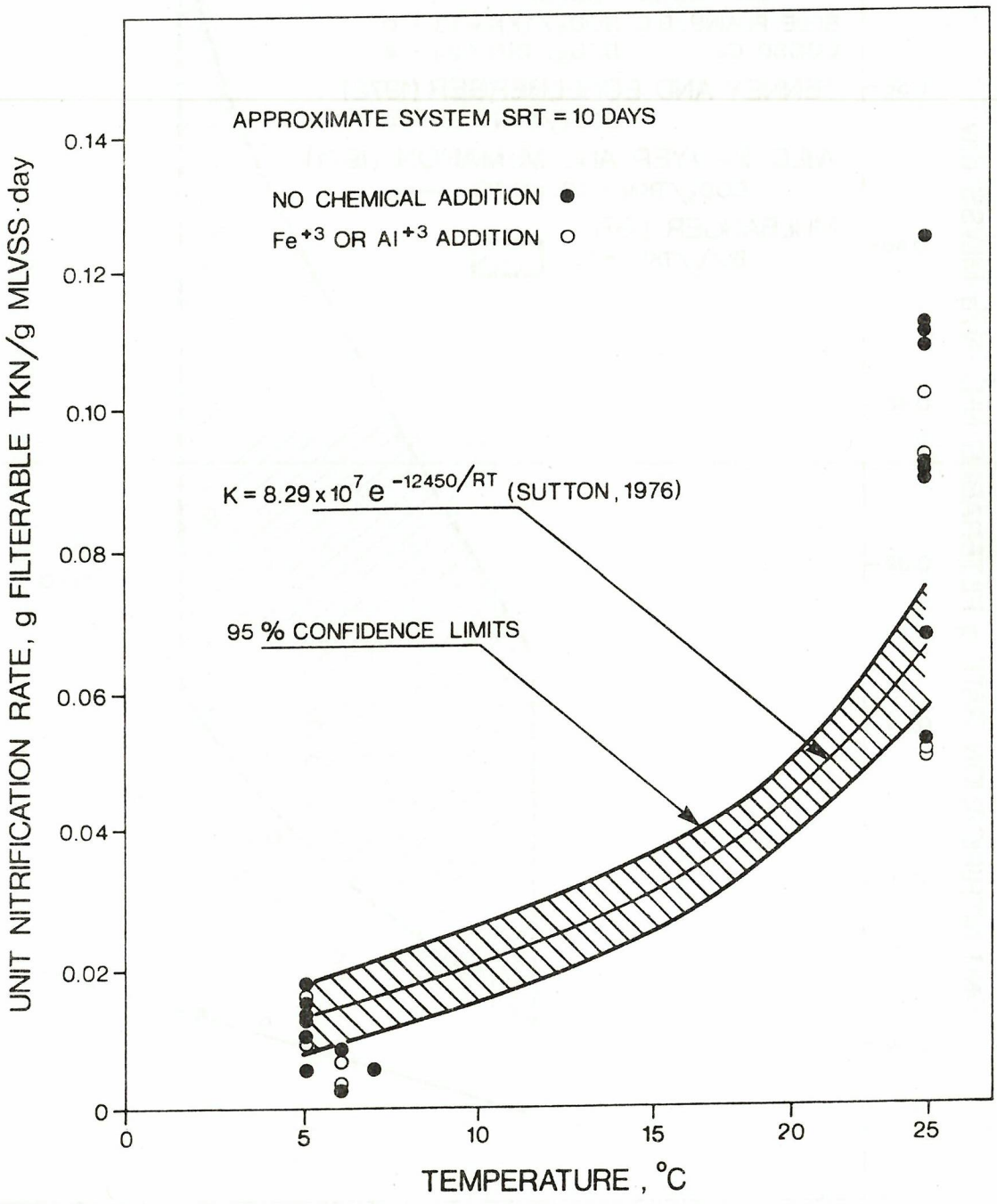


FIGURE 11. TEMPERATURE DEPENDENCY OF THE UNIT RATE OF NITRIFICATION IN COMBINED SLUDGE SYSTEMS

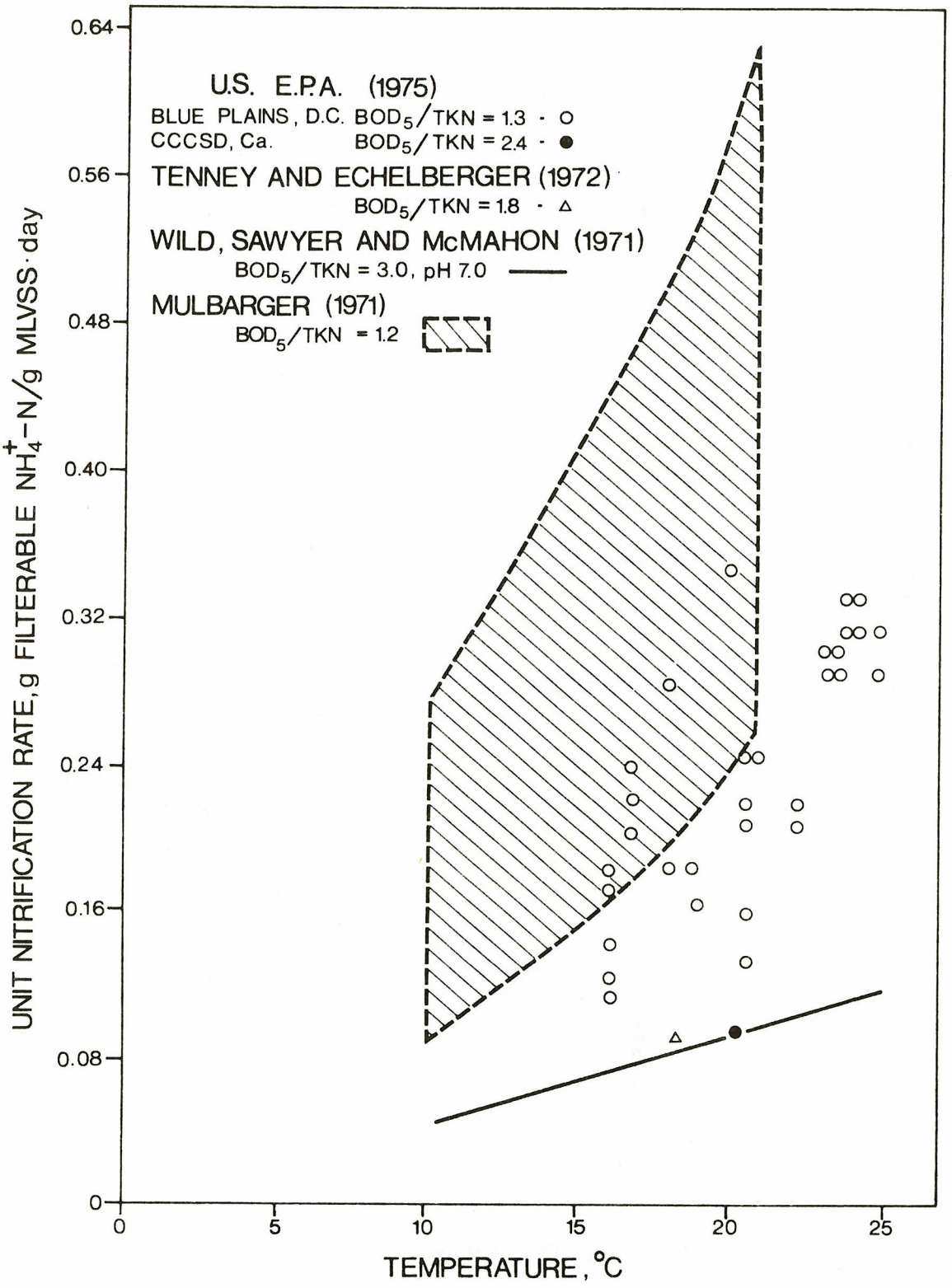


FIGURE 12. NITRIFICATION RATES OBSERVED BY VARIOUS AUTHORS

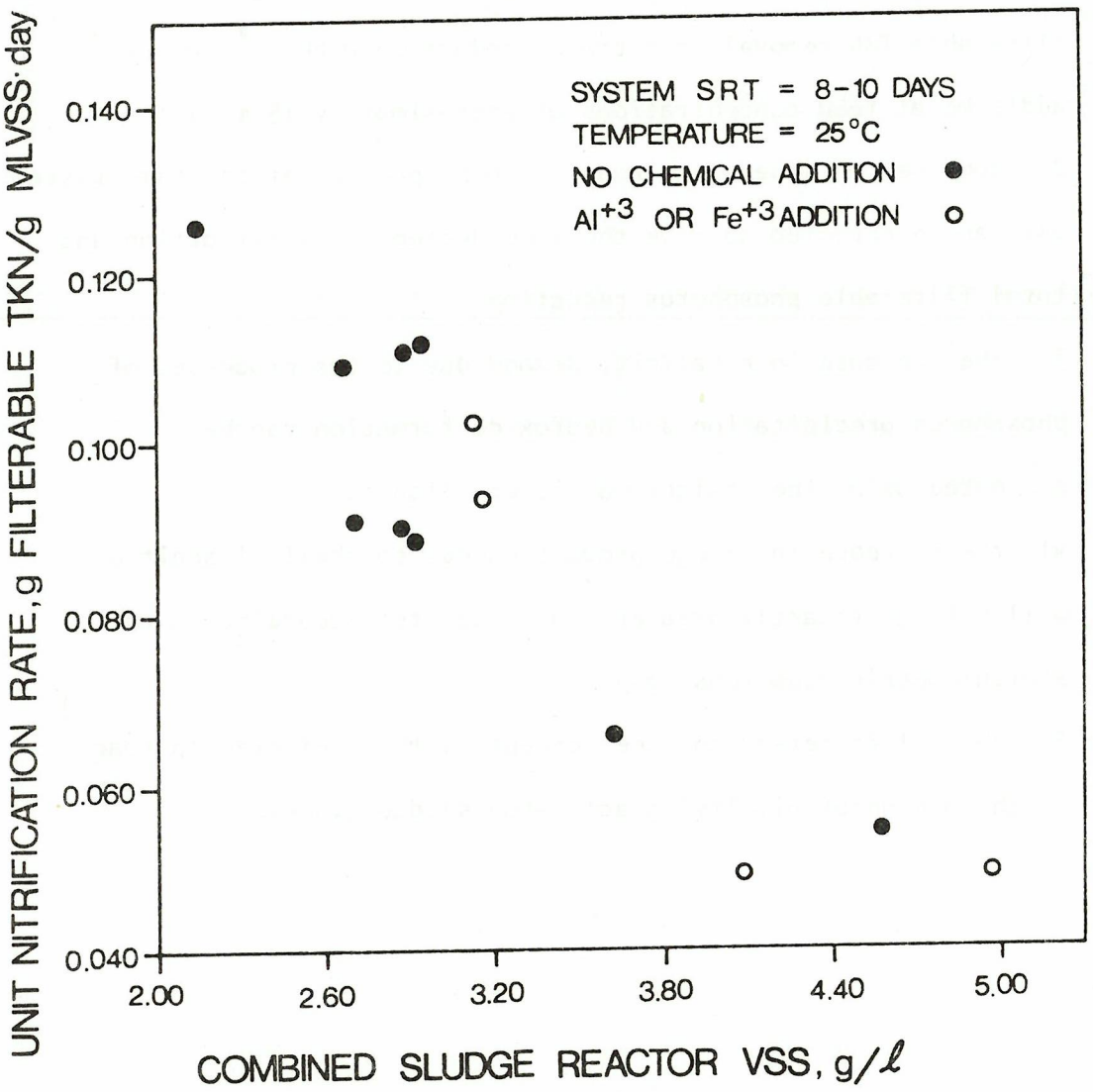


FIGURE 13. NITRIFICATION RATE VARIATION WITH REACTOR VSS AT EQUAL SRT'S

organisms responsible for organic carbon removal necessitates maintaining a higher reactor volatile solids concentration at higher influent organic carbon levels in order to maintain the same SRT.

DESIGN IMPLICATIONS

In designing nitrifying activated sludge systems with integrated phosphorus precipitation the following should be considered:

1) no significant effect can be expected on the degree of filterable TKN removal or nitrate production with Al^{+3} or Fe^{+3} addition at feed concentrations of approximately 15 mg/l,

2) combined or separate sludge systems operated at the same system SRT can be expected to give the same degree of nitrification and total filterable phosphorus reduction,

3) the increase in alkalinity demand due to the processes of phosphorus precipitation and hydroxide formation can be estimated using the stoichiometric equations,

4) the increase in sludge production due to chemical addition will be significantly greater than predicted according to the stoichiometric equations, and

5) the solids retention time concept is the preferred approach in the design of nitrifying activated sludge systems.

REFERENCES

Black, S.A., "Phosphorus Removal by Chemical Addition to an Extended Aeration Activated Sludge Plant", Ministry of the Environment, Ontario, Research Publication No. 39, (1971).

Hais, A.B., J.B. Stamberg and D.F. Bishop, "Alum Addition to Activated Sludge with Tertiary Solids Removal", Chem. Eng. Prog. Symp. Series, 68, 124,35, (1976).

Lawrence, A.W. and C.G. Brown, "Design and Control of Nitrifying Activated Sludge Systems", J. Wat. Pollut. Control Fed., 48, 7, 1779, (1976).

Lawrence, A.W. and P.L. McCarty, "United Basis for Biological Treatment Design and Operation", J. Sanitary Eng. Division ASCE, 96, SA3, 757, (1970).

Long, D.A., J.B. Nesbitt and R.R. Kountz, "Soluble Phosphate Removal in the Activated Sludge Process - A Two Year Plant Scale Study", Proc. 26th Indust. Waste Conf., Purdue University (1971).

Mulbarger, M.C., "Nitrification and Denitrification in Activated Sludge Systems", J. Wat. Pollut. Control Fed., 43, 10, 2059 (1971).

Murphy, K.L., "Significance of Flow Patterns and Mixing in Biological Waste Treatment", Biological Waste Treatment, edited by R.P. Canale, Interscience Publishers, 35, (1971).

Snedecor, G.W. and W.G. Cochran, "Statistical Methods", Iowa State Univ. Press, Ames, Iowa, (1967).

Sutton, P.M., "Steady-State and Dynamic Behaviour of Combined and Separate Sludge Carbon Removal-Nitrification Systems", Ph.D. Thesis, McMaster University, Hamilton, Canada, (1976).

Sutton, P.M., K.L. Murphy, B.E. Jank and B.A. Monaghan, "Reliability of Nitrification Systems with Integrated Phosphorus Precipitation", Canada-Ontario Agreement Research Report No. 64, 1977.

Sutton, P.M., K.L. Murphy, B.E. Jank and B.A. Monaghan, "Efficacy of Biological Nitrification", J. Wat. Pollut. Control Fed., 47, 11, 2665 (1975).

Tenney, M.W., and W.F. Echelberger, "Removal of Organic and Eutrophying Pollutants by Chemical-Biological Treatment", EPA Report No. R2-72-076, (1972).

Unz, R.F. and J.A. Davis, "Microbiology of Combined Chemical-Biological Treatment", J. Wat. Pollut. Control Fed., 47, 185, (1975).

U.S. EPA, "Process Design Manual for Nitrogen Control", Office of Technology Transfer, Washington, D.C., (1975).

U.S. EPA, "Process Design Manual for Phosphorus Removal", Office of Technology Transfer, Washington, D.C., (1976).

Wild, H.E., C.N. Sawyer and T.C. McMahon, "Factors Affecting Nitrification Kinetics", J. Wat. Pollut. Control Fed., 43, 1845, (1971).

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ORGANIC CHEMICALS INDUSTRY
NITRIFICATION STUDY

by

T.R. Bridle

Presented at
NUTRIENT CONTROL TECHNOLOGY SEMINAR
Calgary, Alberta
Feburary 7-8, 1980

ABSTRACT

Operating conditions were defined for Du Pont of Canada Ltd.'s single sludge nitrification-denitrification system at Maitland, Ontario. The plant treats a complex organic chemical industry wastewater containing high concentrations of organic carbon, and organic and inorganic nitrogen.

Bench scale studies demonstrated that a minimum aerobic SRT of 25 to 30 days was required to achieve complete nitrification at an aeration basin temperature of 20°C and 55 to 60 days SRT at 10°C.

In full scale tests, under summer operating conditions, complete nitrification was demonstrated and effluent exhibited zero fish mortality. During the winter, at an equilibrium SRT in excess of 60 days, the process was unable to maintain nitrification due to the sensitivity of nitrifiers to variations in feed composition and temperature.

NITRIFICATION AND DENITRIFICATION OF AN INDUSTRIAL WASTEWATER: RESEARCH AND DEMONSTRATION STUDIES

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Juerg Arnold, Du Pont of Canada Ltd., Maitland, Ontario.

INTRODUCTION

Effluent regulations and guidelines are currently being developed as part of a Canadian federal government program to control pollution from the industrial community. The organic chemical industry is one industrial sector for which the controls are being developed under the Fisheries Act of Canada. The controls are developed through a consultative approach between industry, and both the federal and provincial governments.

Du Pont of Canada Ltd. has played an active role in the development of the regulations for the organic chemical industry. Officials from the company participated in task force activities and their nitrification-denitrification waste treatment facility at Maitland, Ontario, provided information on the capability of a biological system to treat a chemical processing plant waste containing high concentrations of organic carbon, and organic and inorganic nitrogen.

One control parameter which will be included in the effluent regulations and guidelines is an acute toxicity limit using rainbow trout. This parameter indirectly necessitates the need for a high degree of ammonia conversion (nitrification) in any biological wastewater treatment system.

The Maitland Works waste treatment facility (Figure 1) consists of a two-stage combined sludge pre-denitrification-nitrification system.

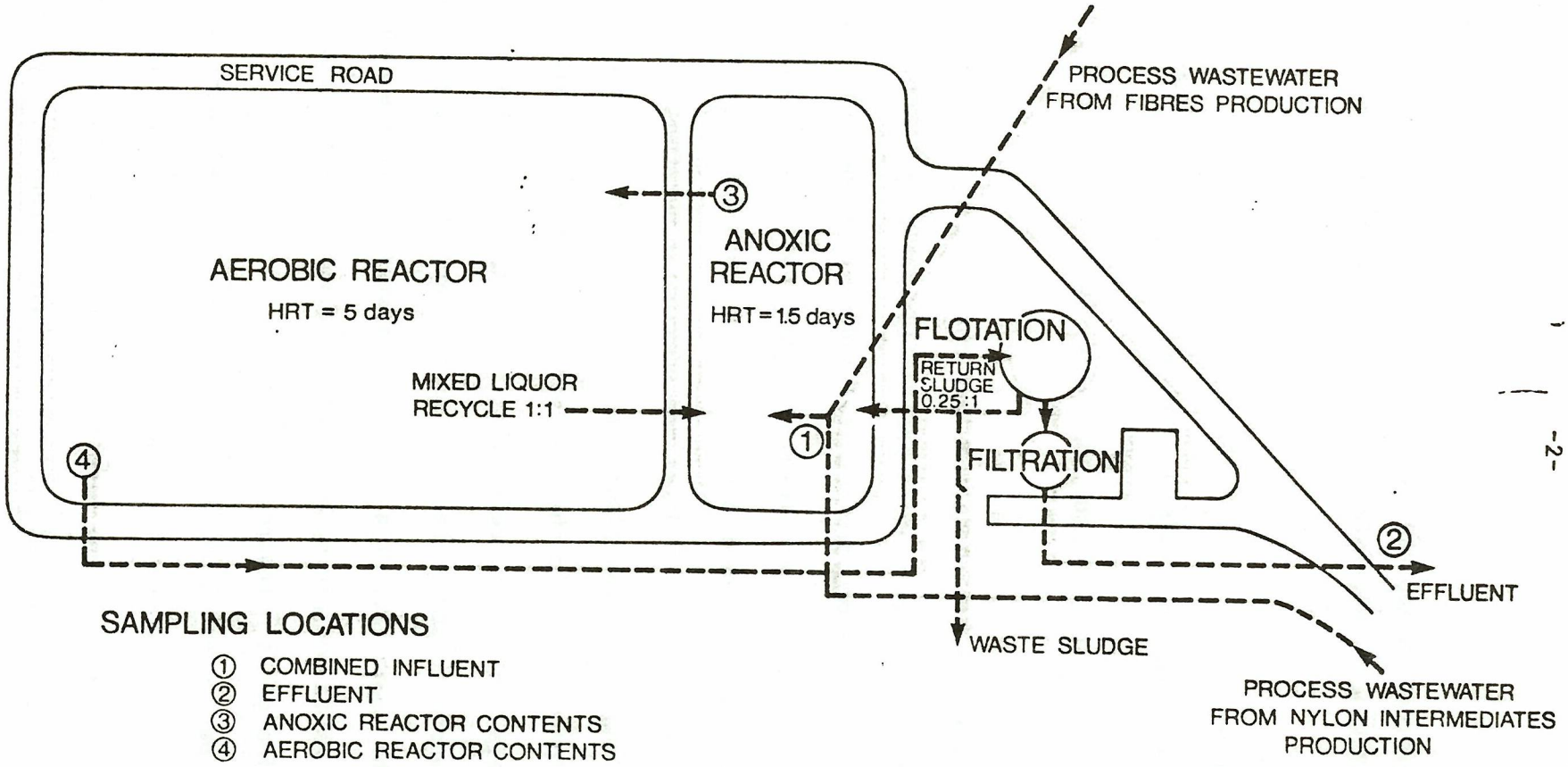


FIGURE 1. DU PONT MAITLAND WASTE TREATMENT FACILITY

Denitrification is accomplished in the anoxic (absence of molecular oxygen) first stage with the organics present in the wastewater providing the external carbon and energy source for nitrate reduction. The conversion of ammonia to nitrate, or nitrification, and oxidation of any biodegradable organics not removed in the first stage, is accomplished in the aerobic second stage. A detailed review of the principles of biological nitrification and denitrification can be found elsewhere[1]. An air flotation unit provides solid-liquid separation at the Maitland plant and a dual media gravity filter is used for final effluent polishing.

Since start-up of the Du Pont Maitland waste treatment plant in 1973, organic carbon removal and denitrification had been excellent while the degree of nitrification had been seasonal with loss of nitrification each winter as the aerobic basin temperature approached 10°C.

In an effort to better define the conditions necessary to obtain nitrification using the Maitland Works process configuration, and to assess the capability of the existing full scale system, a program involving bench scale and full scale demonstration studies was carried out over a period of two years involving personnel from Du Pont of Canada Ltd. and from the Wastewater Technology Centre in Burlington, Ontario. In the first phase of the program -

(1) bench scale units were operated at the Wastewater Technology Centre to establish the minimum solids retention time (SRT) for nitrification and denitrification under mixed liquor temperature conditions of approximately 20°C, and

(2) full scale plant demonstration studies were carried out involving two periods of controlled operation under high and low temperature conditions.

From the results of the first phase it was concluded that additional bench scale work was necessary to -

(1) assess the SRT conditions necessary to obtain nitrification using the existing full scale process configuration at mixed liquor temperatures down to 10°C, and

(2) assess the capabilities of alternative process configurations to achieve nitrification at temperatures down to 10°C.

BENCH SCALE AND FULL SCALE DEMONSTRATION STUDIES

Process Facilities and Operation

Du Pont of Canada Ltd. manufactures nylon intermediates, acrylic fibre and other chemicals at the Maitland Works located on the St. Lawrence River, 100 miles west of Montreal. Wastewater, principally from nylon intermediates manufacture contains non-recoverable nitric and nitrous acid, organic carbon, mainly C_1-C_5 monobasic acids, ammonia and organic nitrogen compounds. A detailed description of the wastewater treatment facility together with a documentation of the problems encountered during the initial start-up have been presented previously[2].

The full scale studies in phase 1 involved operation of the waste treatment plant (Figure 1) under both high and low temperature conditions. The total flow rate to the plant was maintained at approximately 11 x 10^3 m^3/d (2.4 x 10^6 Igal/day). The maximum to average to minimum flow variation over the day was approximately 1.1:1.0:0.9. The waste concentration and variation to the plant during the high temperature period (Figure 6) is given in Table 1. The waste concentration was decreased by approximately 25 percent during the low temperature period (Figure 6) by by-passing a portion of one of the process wastewater streams. The characteristics and

TABLE 1. COMBINED PROCESS WASTEWATER CHARACTERISTICS

Characteristic†	Phase 1 Levels		Phase 2 Levels	
	Median	90%*	Median	90%*
BOD_5 (unacclimated seed)	1230	1870	-	-
FOC (filtered organic carbon)	800	1250	1115	1325
Filtered COD	3000	4000	3433	4260
Filtered TKN	190	300	157	207
NH_4-N	130	225	96	130
NO_3-N + NO_2-N	175	260	206	287
SS	8	20	-	-
pH	4.5	6.5	3.4	4.6

NOTE: * 90% of observations were equal to or less than stated value

† all values in mg/ℓ except pH

variability of the process wastewater obtained from Du Pont Maitland for the phase 1 and 2 bench scale studies are given in Table 1.

Since a prime goal in the demonstration study was to assess the capability of the full scale system to achieve nitrification, particular attention was paid to maintaining the dissolved oxygen concentration throughout the aerobic reactor above a minimum level (2.0 mg/ℓ) and controlling the SRT.

Twenty-four hour composite sampling was carried out seven times per week on the influent stream and five times per week on the effluent stream. Various analyses were performed on samples taken from the influent and effluent sampling locations (Figure 1, locations 1 and 2). Grab samples from the anoxic and aerobic reactors (Figure 1, locations 3 and 4) were taken five times per week.

Fish bioassay tests were carried out to assess the toxicity of the effluent. The tests involved 96-hour static bioassays using juvenile rainbow trout (Salmo gairdneri). Ten fish averaging 7.5 cm in length were exposed to various sample dilutions. The dissolved oxygen was maintained at 8.0 mg/ℓ in the sample containers. The pH of the samples was not altered but the temperatures were maintained at 15°C. Observations on mortality were made at 0.25, 0.5, 1, 2, 4, 8, 18, and 24 hours and every 24 hours thereafter for the duration of the 96-hour test.

The bench scale studies conducted at the Wastewater Technology Centre in phase 1 prior to the full scale demonstration studies involved operation of three parallel process configurations. Each unit (Figure 2) consisted of a first stage complete mix anoxic stirred tank reactor (D1) followed by a second stage complete mix aerobic reactor (D2). An upflow clarifier provided solid-liquid separation. Each configuration was operated under

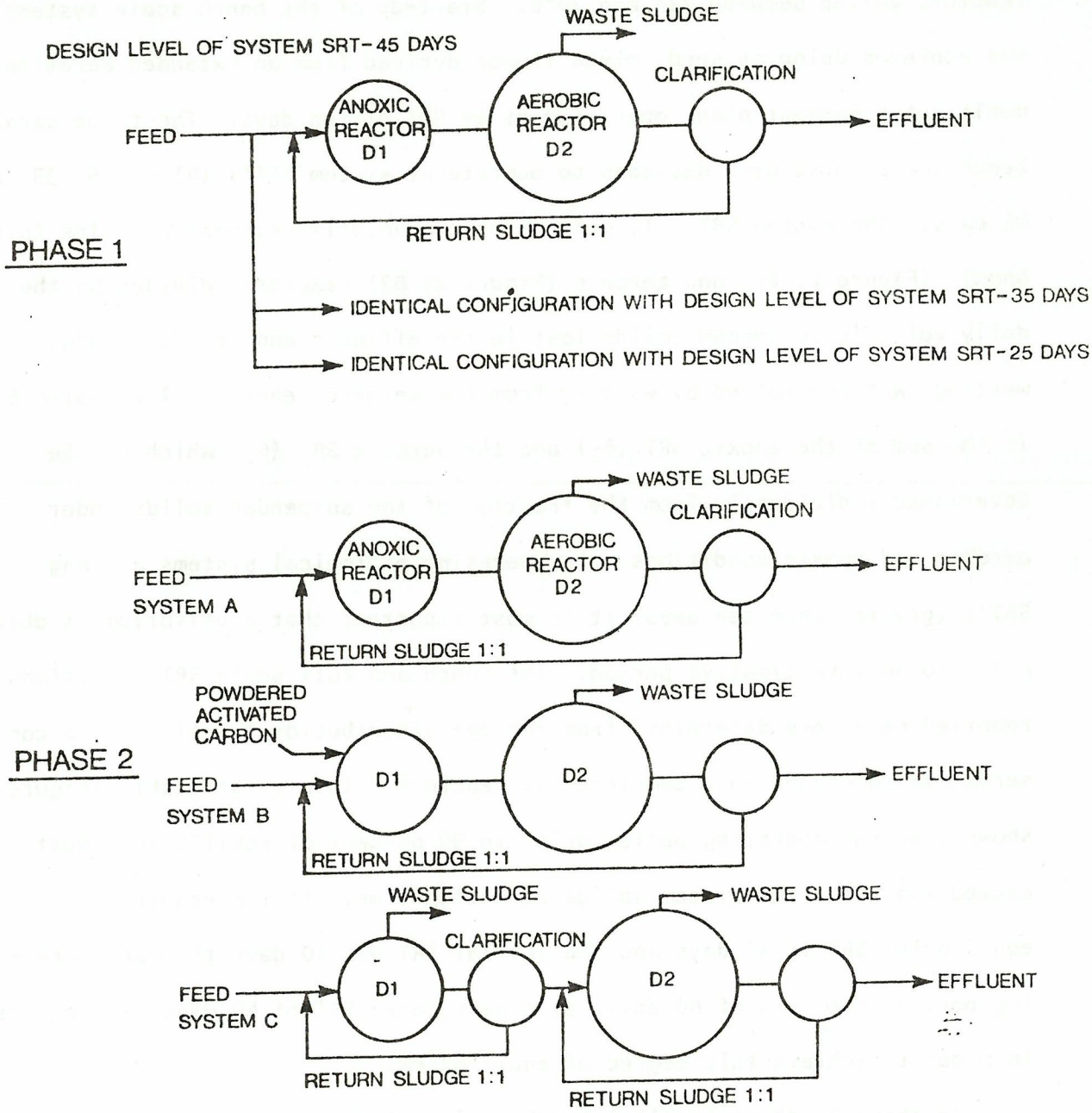


FIGURE 2. PHASE 1 AND 2 BENCH SCALE DENITRIFICATION-NITRIFICATION CONFIGURATIONS

identical flow rate, temperature, and pH conditions but at three different levels of sysem solids retention time (SRT). The temperature in the reactors varied between 22° and 24°C. Start-up of the bench scale systems was achieved using as seed, mixed liquor derived from an extended aeration municipal treatment plant operating at an SRT of ten days. The three parallel bench scale units were designed to operate at system SRT's (θ) of 25, 35, and 45 days. The system SRT (θ), defined as the volatile suspended solids in the anoxic (Figure 2, D1) and aerobic (Figure 2, D2) reactors, divided by the daily volatile suspended solids lost in the effluent and through sludge wasting, was controlled by wasting from the aerobic reactor. The system SRT is the sum of the anoxic SRT (θ_1) and the aerobic SRT (θ_2) which can be determined individually from the fraction of the suspended solids under aerobic and anoxic conditions. In operating biological systems at long SRT's (greater than ten days) it is most important that equilibrium is obtained prior to an investigative period. The bench and full scale SRT conditions reported here were determined from the age distribution function for a conservative substance in a complete mix reactor[3]. This relationship (Figure 3) shows that the operating period prior to 90 percent of equilibrium, must exceed two times the system solids retention time. If the required equilibrium SRT is 40 days and the initial SRT was 10 days then an operating period in excess of 60 days, at a calculated SRT of 40 days, is required in order to achieve this degree of equilibrium.

In phase 2, three parallel bench scale systems were operated with three different process configurations (Figure 2). The configurations were:

(1) a two-stage single sludge pre-denitrification-nitrification system (system A),

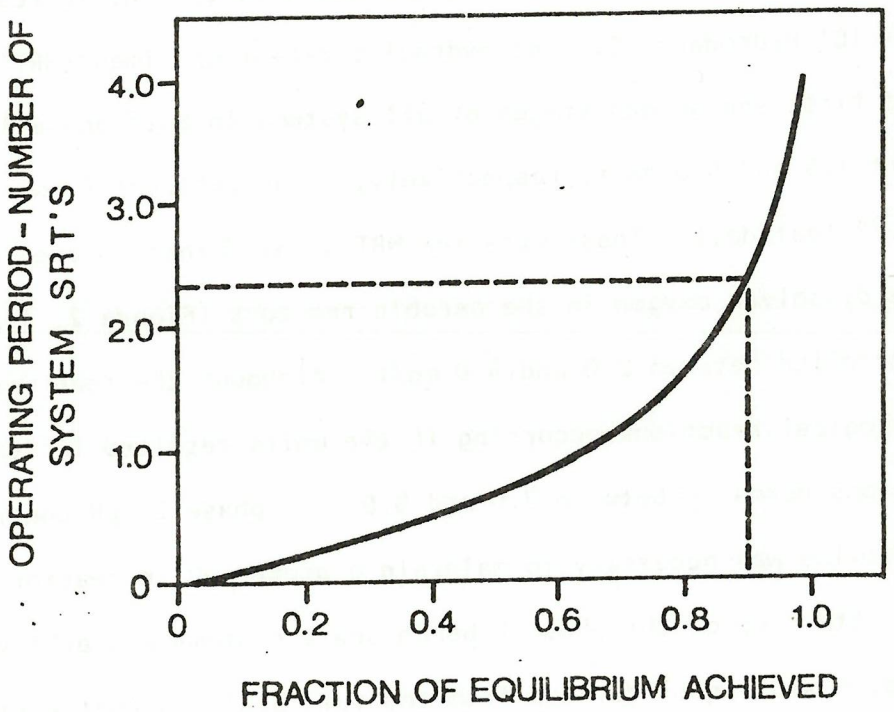


FIGURE 3. THEORETICAL RELATIONSHIP BETWEEN EQUILIBRIUM AND PERIOD OF OPERATION

(2) a two-stage single sludge pre-denitrification-nitrification system (system B) with 50 mg/ℓ of powdered activated carbon (PAC) addition to the anoxic reactor (Figure 2), and

(3) a two-stage separate sludge pre-denitrification-nitrification system (system C).

The powdered activated carbon added to the anoxic reactor of system B was ICI Hydrodarco C. The hydraulic retention times (HRT's) maintained in the first and second stages of all systems in both phase 1 and phase 2, were 1.5 and 5.0 days, respectively, at an influent flow rate of 3.73 ℓ/d (0.82 lgal/day). These were the HRT's maintained in the full scale facility. The dissolved oxygen in the aerobic reactors (Figure 2, reactor D2) was controlled between 2.0 and 4.0 mg/ℓ. Although the feed was acidic the biological reactions occurring in the units resulted in reactor pH conditions normally between 7.0 and 8.0. In phase 2, pH control with sodium hydroxide was necessary to maintain a proper pH in reactor D1 of system C.

Start-up of the phase 2 bench scale systems was achieved using as seed, mixed liquor derived from the full scale Du Pont waste treatment plant operating at a system SRT of 40 days. The bench scale units were initially operated at approximately 18°C while the equilibrium system SRT in each configuration was increased to 80 days. During this initial acclimation period PAC was added to system B at a rate of 150 mg/ℓ to decrease the time necessary to approach the equilibrium concentration in the system of approximately 700 mg/ℓ. This is the equilibrium which will be achieved at the specified HRT conditions when the system is operated at an 80 day SRT and PAC is added at a rate of 50 mg/ℓ. Once equilibrium SRT conditions were achieved in configurations A, B, and C and once the equilibrium PAC concentration was achieved in system B, the temperature in

all systems was reduced to approximately 10°C at a rate of 2°C per week. A controlled environmental laboratory allowed maintenance of the temperature at 10°C while the equilibrium SRT was varied in the systems from 60 to 80 days (Figure 10).

Following assessment of the performance of the systems under these conditions alterations were made to systems B and C. A submersible heater was placed in reactor D1, system B, and the temperature increased to approximately 17°C at a rate of 3.5°C per day. System C was converted to a configuration identical to system A and then adjusted to an equilibrium system SRT of 60 days. At this point a submersible heater was placed in reactor D1 and the temperature adjusted to approximately 17°C as in system B. The temperature in system A and in reactor D2 of systems B and C was maintained at approximately 10°C. The purpose of creating a temperature difference between reactors D1 and D2 of systems B and C, was to simulate conditions observed in the full scale facility during the low temperature period (Figure 6).

The performance of the bench scale units was assessed by sampling on the average of two times per week. Influent grab samples and effluent composite samples represented weekly composite results. The samples were analyzed according to procedures given elsewhere[4]. Mixed liquor grab samples were taken once per day for five days per week for volatile suspended solids determination and SRT control.

RESULTS AND DISCUSSION

In the bench and full scale nitrification and denitrification configurations studied here, the denitrifying organisms reduced nitrate to dinitrogen (N_2) using the organics present in the wastewater as the carbon and energy source, and the nitrifying organisms oxidized ammonia to nitrate.

The rate limiting step in the nitrification reaction is the conversion of ammonia to nitrite by the Nitrosomonas (NS) organisms. The growth rate of Nitrosomonas under steady-state conditions can be described by:

$$U_N = U_N^* \frac{N}{K_N + N} \qquad (1)$$

where

U_N = growth rate of NS (d^{-1}),

U_N^* = maximum growth rate of NS (d^{-1}),

N = NH_4-N concentration (mg/ℓ), and

K_N = N concentration at which $U_N = 0.5\ U_N^*$ (mg/ℓ).

Equation 1 indicates a zero order relationship between the oxidation rate and substrate concentration when the value of K_N is much less than N. The zero-order nature of the nitrification reaction down to low ammonia-nitrogen concentrations has been reported by numerous authors treating both municipal and industrial wastewaters[5,6,7].

The oxidation of ammonia can be related to the NS growth rate according to:

$$\frac{1}{X_V}\frac{DN}{dt} = q_N = \frac{U_N}{Y_N} = q_N^* \frac{N}{K_N + N} \qquad (2)$$

where

X_V = NS mixed liquor volatile suspended solids concentration (mg/ℓ),

q_N = ammonia oxidation rate (mg NH_4-N oxidized/mg NS VSS·d),

q_N^* = maximum ammonia oxidation rate (mg NH_4-N oxidized/mg NS VSS·d), and

Y_N = organism net yield coefficient (mg NS VSS grown/mg NH_4-N oxidized).

The growth rate of Nitrosomonas and all organisms is fixed according to the solids retention time (SRT) or sludge age:

$$SRT = \frac{1}{U_N} \qquad (3)$$

In the process configurations studied here, since nitrifiers require elemental oxygen for growth, the SRT in equation 3 may be expected to be the aerobic SRT (θ_2). In combined sludge aerobic-anoxic nitrification and denitrification systems treating municipal wastewaters, it has been found that the degree of nitrification was dependent on the aerobic SRT[8,9].

Phase 1: Bench Scale and Full Scale Studies

The zero-order nature of the nitrification reaction was demonstrated at the completion of the phase 1 bench scale studies from the results (Figure 4) of batch studies in the aerobic reactors achieving nitrification. The ammonia oxidation rate, based on total volatile suspended solids, determined from Figure 4 is approximately one-tenth of the rate observed in carbon oxidation-nitrification systems treating municipal wastewaters. Therefore, from the growth rate and ammonia oxidation rate relationship (equation 2), the aerobic SRT required to obtain nitrification may be expected to be approximately ten times greater than that required for systems treating municipal wastewaters. The continuous bench scale study results indicated a minimum aerobic SRT of 25 days was required to obtain complete nitrification at approximately 22° to 24°C (Figure 5). The minimum SRT requirements in treating municipal wastewaters at this temperature level is approximately two to three days[1,9].

The minimum aerobic SRT at which the nitrification performance was assessed was 17 days (Figure 5). This corresponded to a system SRT of 22 days or an anoxic reactor SRT (θ_1) of five days. Although at the lower

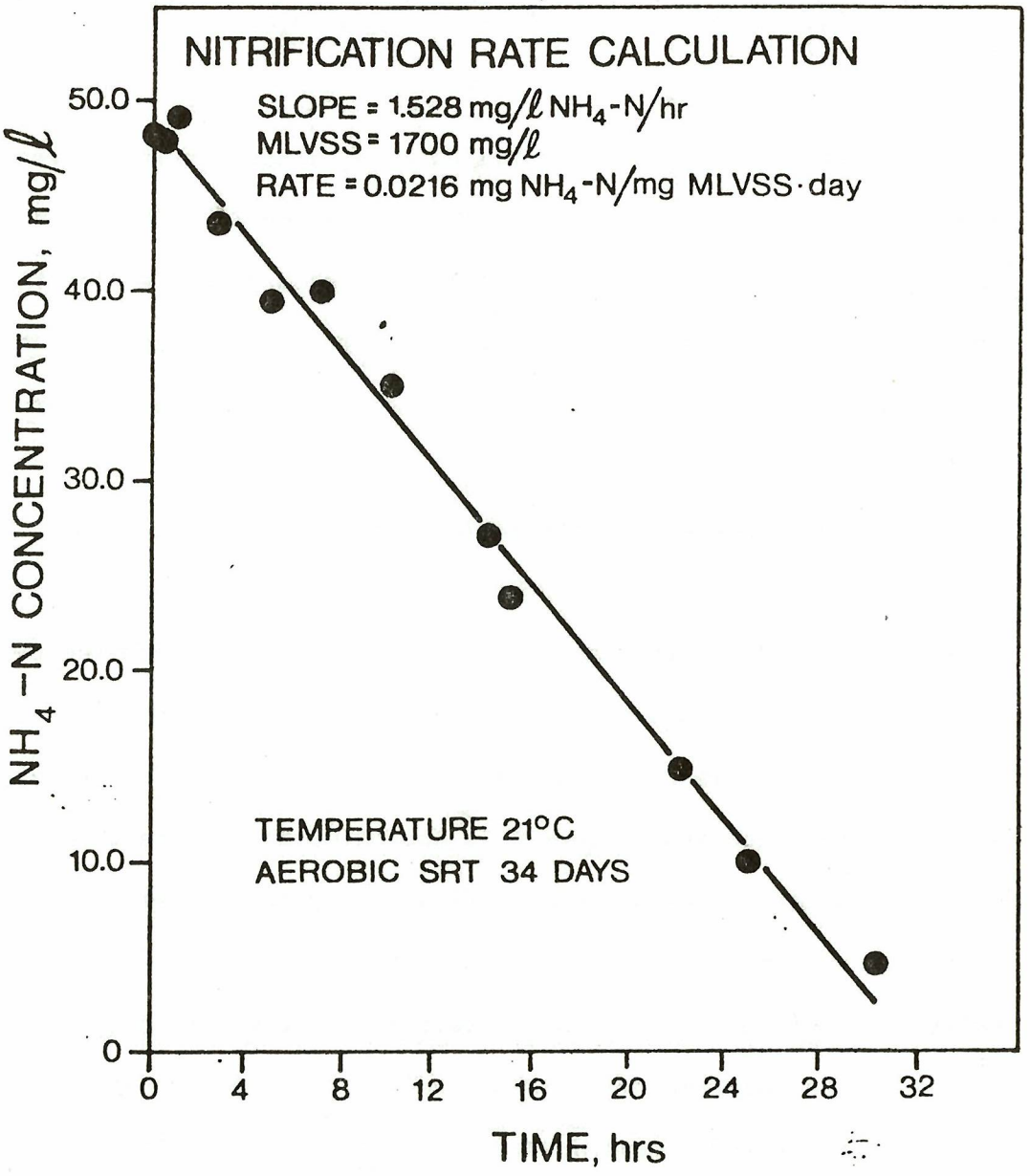


FIGURE 4. BATCH NITRIFICATION RESULTS

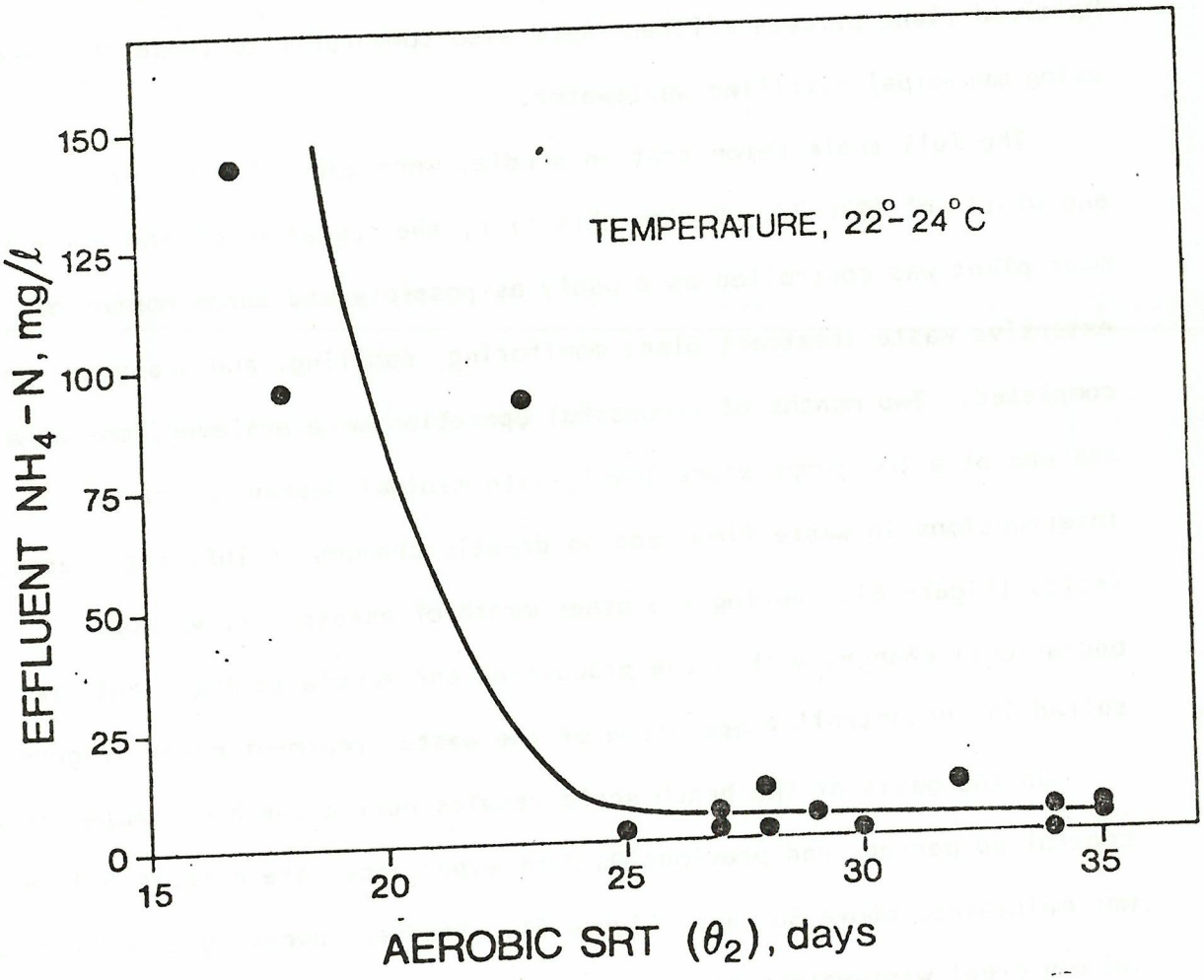


FIGURE 5. BENCH SCALE NITRIFICATION PERFORMANCE

SRT conditions, ammonia did appear in the effluent from the aerobic reactor, complete denitrification was observed in the anoxic reactor (NO_3-N + NO_2-N less than 2.0 mg/ℓ). Denitrification of municipal nitrified wastewater under non-carbon limiting conditions has been achieved in combined and separate sludge systems at comparable anoxic SRT's[1,9]. The denitrification rates determined from previous studies[2] using the Du Pont Maitland plant process effluent were also comparable to those observed using municipal nitrified wastewater.

The full scale demonstration studies were carried out over the fall and winter of 1976-77. During this time, the operation of the waste treatment plant was controlled as closely as possible and three months of extensive waste treatment plant monitoring, sampling, and analyses, were completed. Two months of successful operation were achieved, one at a high and one at a low temperature level, with minimal mechanical failures, few interruptions in waste flow, and no drastic changes in influent characteristics (Figure 6). During the other month of assessment, various operational changes within the processing and manufacturing plants resulted in uncontrolled operation of the waste treatment plant (Figure 6).

On the basis of the bench scale results during the high temperature controlled period, and previous Du Pont experience, the aerobic SRT (θ_2) was maintained above 30 days (Figure 6). Studies involving nitrification of municipal wastewaters have indicated that the nitrification rate approximately doubles for a rise in temperature from 10° to 20°C[1,9]. Previous Du Pont laboratory and full scale studies[2] had indicated the need for high SRT, and with this in mind, SRT was maintained above 60 days in the low temperature controlled period (Figure 6).

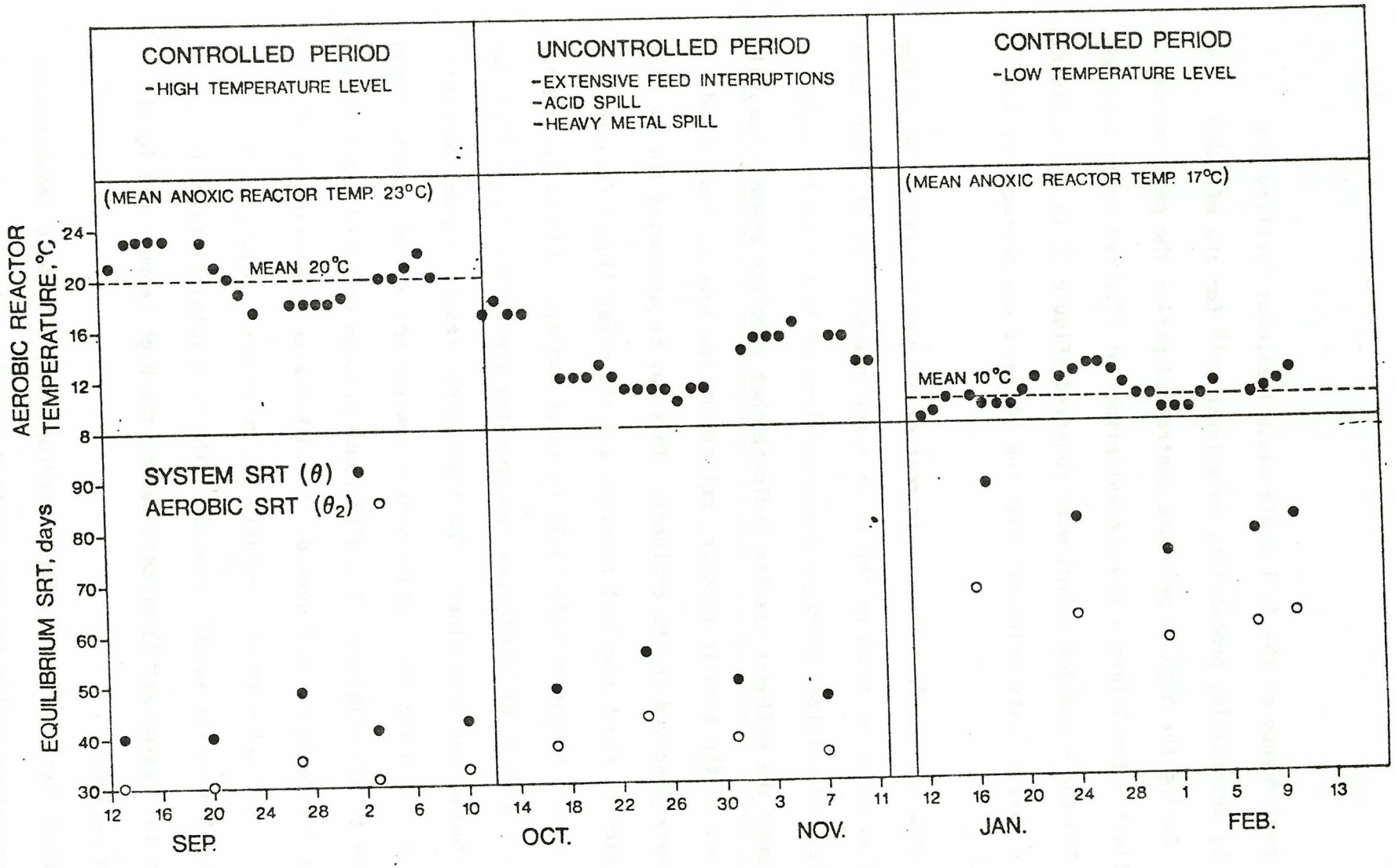


FIGURE 6. FULL SCALE PLANT OPERATION DURING DEMONSTRATION STUDIES

The performance of the full scale waste treatment facility was assessed by determining probability relationships[10] for the effluent results. During the high temperature controlled period the performance was excellent. Consistently low concentrations of filtered organic carbon, filtered TKN and suspended solids were observed (Figure 7) in the effluent. The fish bioassay tests indicated that the effluent was non-acutely lethal (Table 2).

The grab samples from the anoxic reactor allowed determination of the denitrification performance of the first stage reactor. Ninety percent of the samples had oxidized nitrogen concentrations of less than 1.2 mg/ℓ.

Although the effluent results indicate that complete ammonia removal was obtained in the aerobic reactor, relatively low levels of oxidized nitrogen were observed in the effluent. This may be accounted for by the configuration of the biological reactors and the mixed liquor recycle (1:1) and return sludge recycle (0.25:1) rates employed (Figure 1). From a mass balance around the biological reactors the theoretical total filtered nitrogen removal was determined. The theoretical results assume complete denitrification in the first stage anoxic reactor and include the nitrogen removed due to assimilation. The difference between observed and theoretical results is depicted in Figure 8. The difference, amounting to 10 percent of the total nitrogen removal, was considered to be due to denitrification in the aerobic reactor. The mean total filtered nitrogen removal in the system was 93 percent during the high temperature level controlled period.

The feed characteristics, flow interruptions, and rapid reduction of liquid temperature during the uncontrolled period resulted in the loss of nitrification (Figure 6). Fish bioassay tests performed during this period

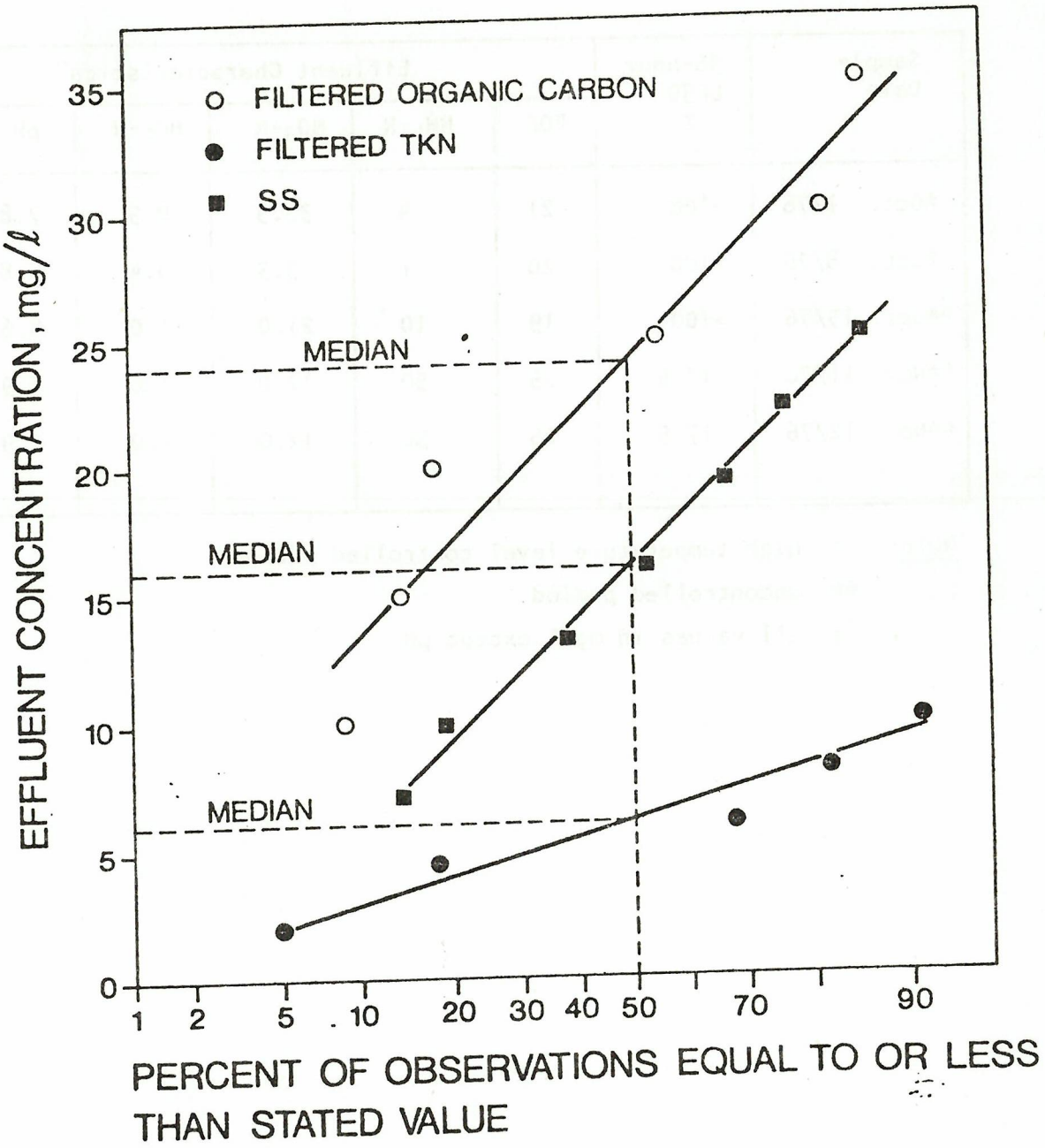


FIGURE 7. EFFLUENT RESULTS DURING HIGH TEMPERATURE LEVEL CONTROLLED PERIOD

TABLE 2. EFFLUENT BIOASSAY RESULTS FROM DU PONT MAITLAND WASTE TREATMENT PLANT

Sample Date	96-hour LC50 %	Effluent Characteristics†				
		FOC	NH_4-N	NO_3-N	NO_2-N	pH
*Oct. 1/76	>100	21	4	33.3	0.9	7.8
*Oct. 8/76	>100	20	1	3.3	0.4	8.8
**Oct. 15/76	>100	19	10	21.0	1.0	8.6
**Nov. 11/76	17.5	25	50	12.0	1.0	7.9
**Nov. 12/76	17.5	25	50	12.0	1.0	7.9

NOTE: * high temperature level controlled period

** uncontrolled period

† all values in mg/ℓ except pH

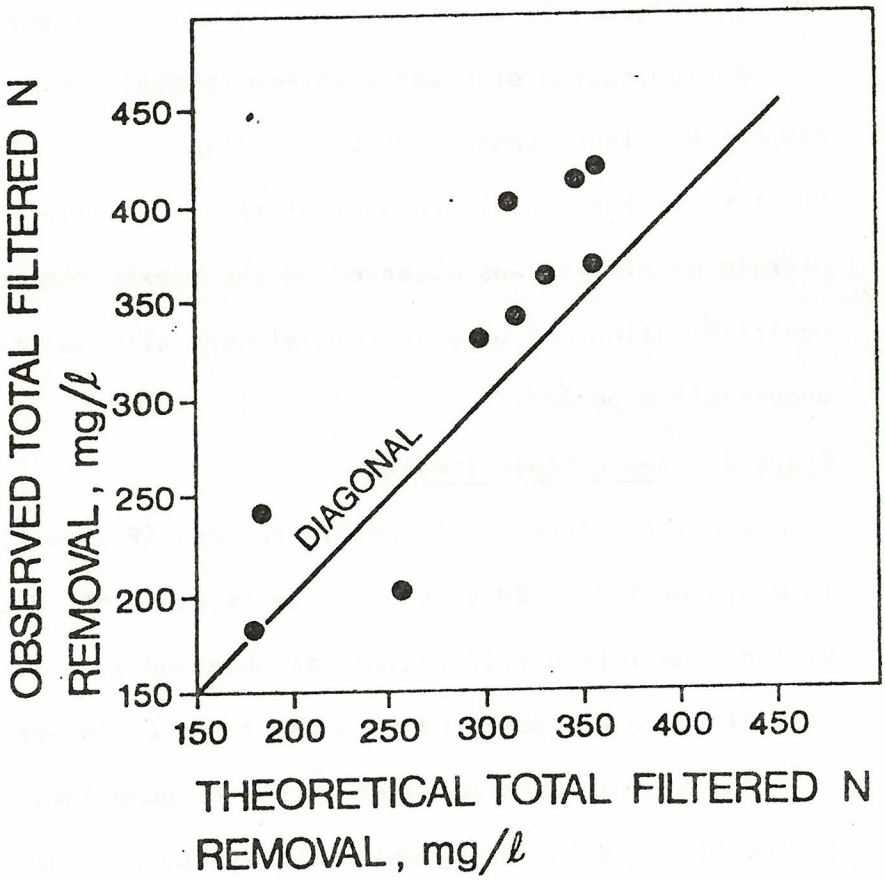


FIGURE 8. OBSERVED AND THEORETICAL N REMOVAL FROM FULL SCALE SYSTEM

indicate the high effluent ammonia concentrations resulted in lethal conditions (Table 2).

Once the manufacturing and processing operations within the production plant had stabilized, the sludge wasting schedule in the treatment plant was altered in preparation for the low temperature assessment period. Even though the resulting equilibrium system SRT achieved was approximately 80 days (Figure 6), nitrification was not re-established.

A high degree of organic carbon removal was maintained during the low temperature level controlled period (Figure 9). Although no nitrification occurred in the aerobic reactor of the full scale system, little or no nitrate or nitrite was observed in the anoxic reactor. High degrees of denitrification and organic removal were also maintained during the uncontrolled period.

Phase 2: Bench Scale Studies

The three bench scale configurations (Figure 2) were first acclimated to a system SRT of 80 days and a temperature of 18°C. Under these conditions complete nitrification was observed in all systems. No loss of nitrification was observed when the temperature was reduced to 10°C.

The systems were operated at the reduced temperature level for a period of 16 weeks during which the system SRT was maintained between 70 and 80 days (Figure 10). This corresponded to an aerobic SRT between 54 and 61.5 days. The performance of the two-stage single sludge systems with and without PAC addition (Figure 2, systems A and B), was excellent with a consistently high degree of nitrification being achieved (Figures 10 and 11). A greater variation in nitrification was observed for the two-stage separate sludge system (Figure 10, system C). Nitrification was measured

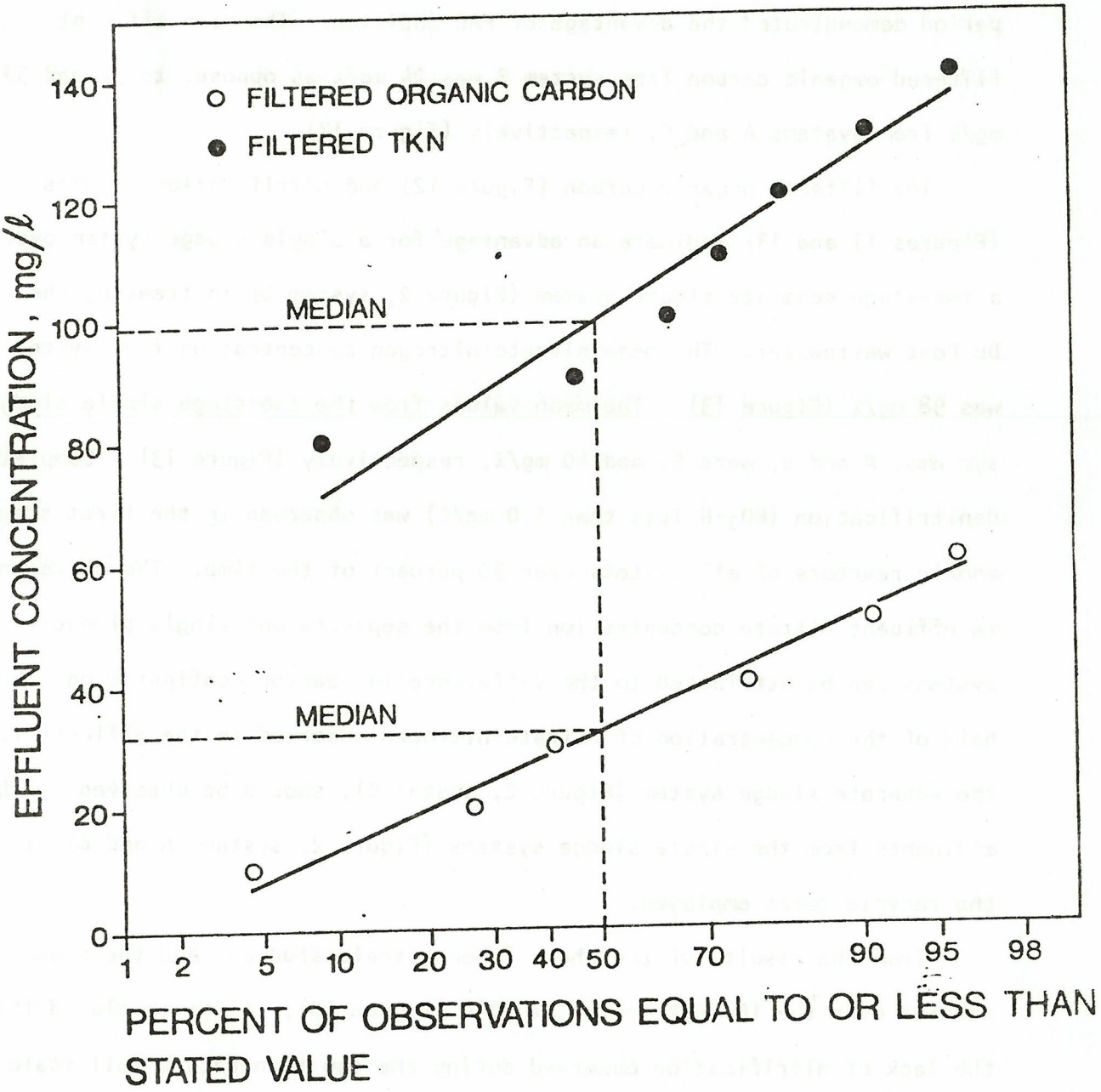


FIGURE 9. EFFLUENT RESULTS DURING LOW TEMPERATURE LEVEL CONTROLLED PERIOD

in terms of the percent filterable TKN removed. At 90 percent filterable TKN removal, the corresponding effluent NH_4-N was less than 5 mg/ℓ.

The effluent filterable organic carbon results over the 16 week period demonstrated the advantage of PAC addition. The mean effluent filtered organic carbon from system B was 24 mg/ℓ as opposed to 32 and 37 mg/ℓ from systems A and C, respectively (Figure 12).

The filtered organic carbon (Figure 12) and nitrification results (Figures 11 and 13) indicate an advantage for a single sludge system over a two-stage separate sludge system (Figure 2, system C) in treating the Du Pont wastewater. The mean nitrate-nitrogen concentration from system C was 98 mg/ℓ (Figure 13). The mean values from the two-stage single sludge systems, A and B, were 50 and 40 mg/ℓ, respectively (Figure 13). Complete denitrification (NO_3-N less than 1.0 mg/ℓ) was observed in the first stage anoxic reactors of all systems over 90 percent of the time. The difference in effluent nitrate concentration from the separate and single sludge systems can be attributed to the difference in reactor configuration. One-half of the concentration of nitrate-nitrogen observed in the effluent from the separate sludge system (Figure 2, system C), should be observed in the effluents from the single sludge systems (Figure 2, systems A and B) at the recycle rates employed.

From the results of the phase 1 bench scale studies, and the phase 2 studies over the 16 week period at 10°C (Figure 10), it was concluded that the lack of nitrification observed during the low temperature full scale demonstration period was due to either -

(1) inability to achieve the aerobic SRT required for nitrifier growth under low temperatures prior to the onset of the low temperature conditions, or

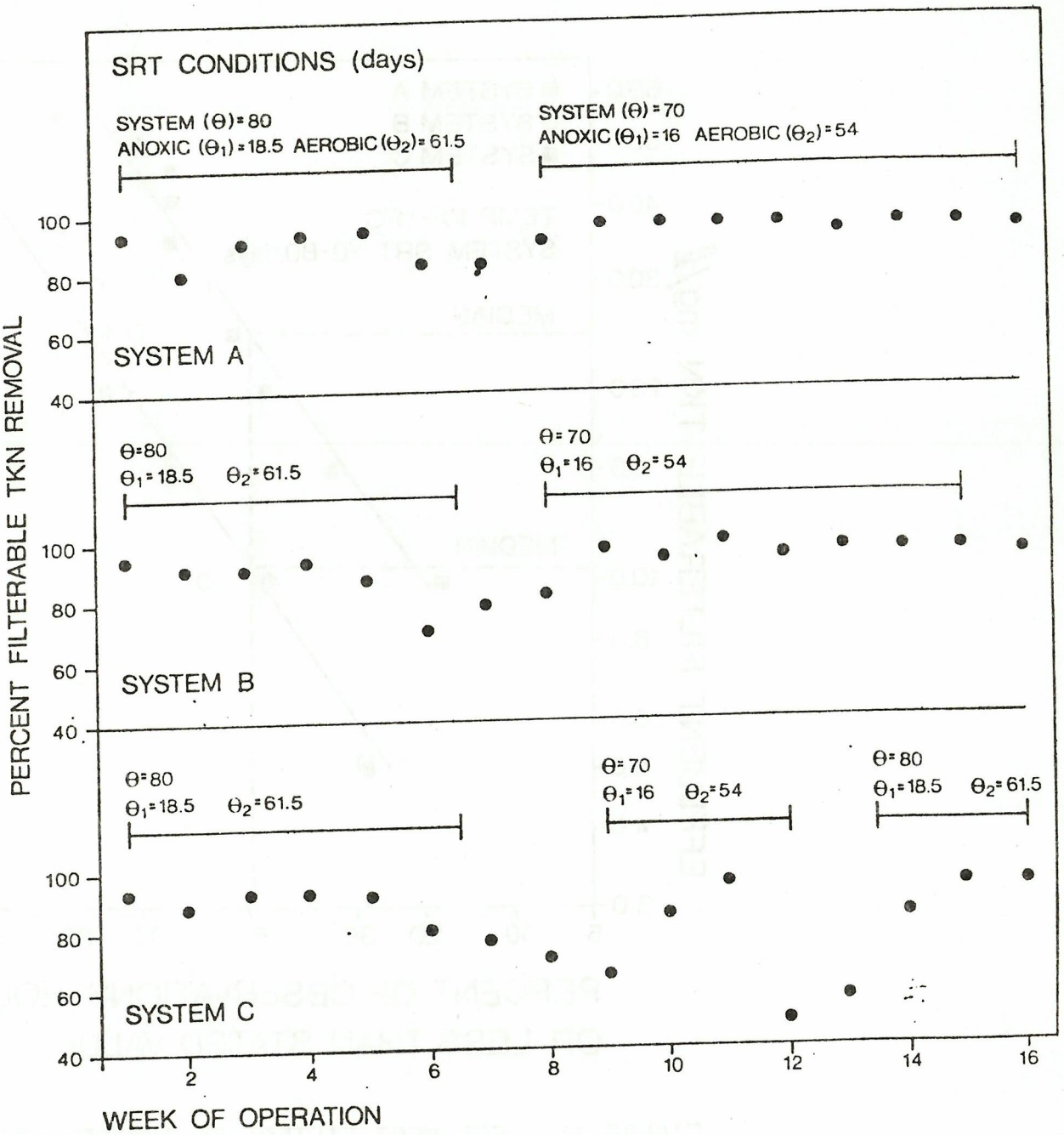


FIGURE 10. PHASE 2 BENCH SCALE STUDIES AT 10° TO 11°C

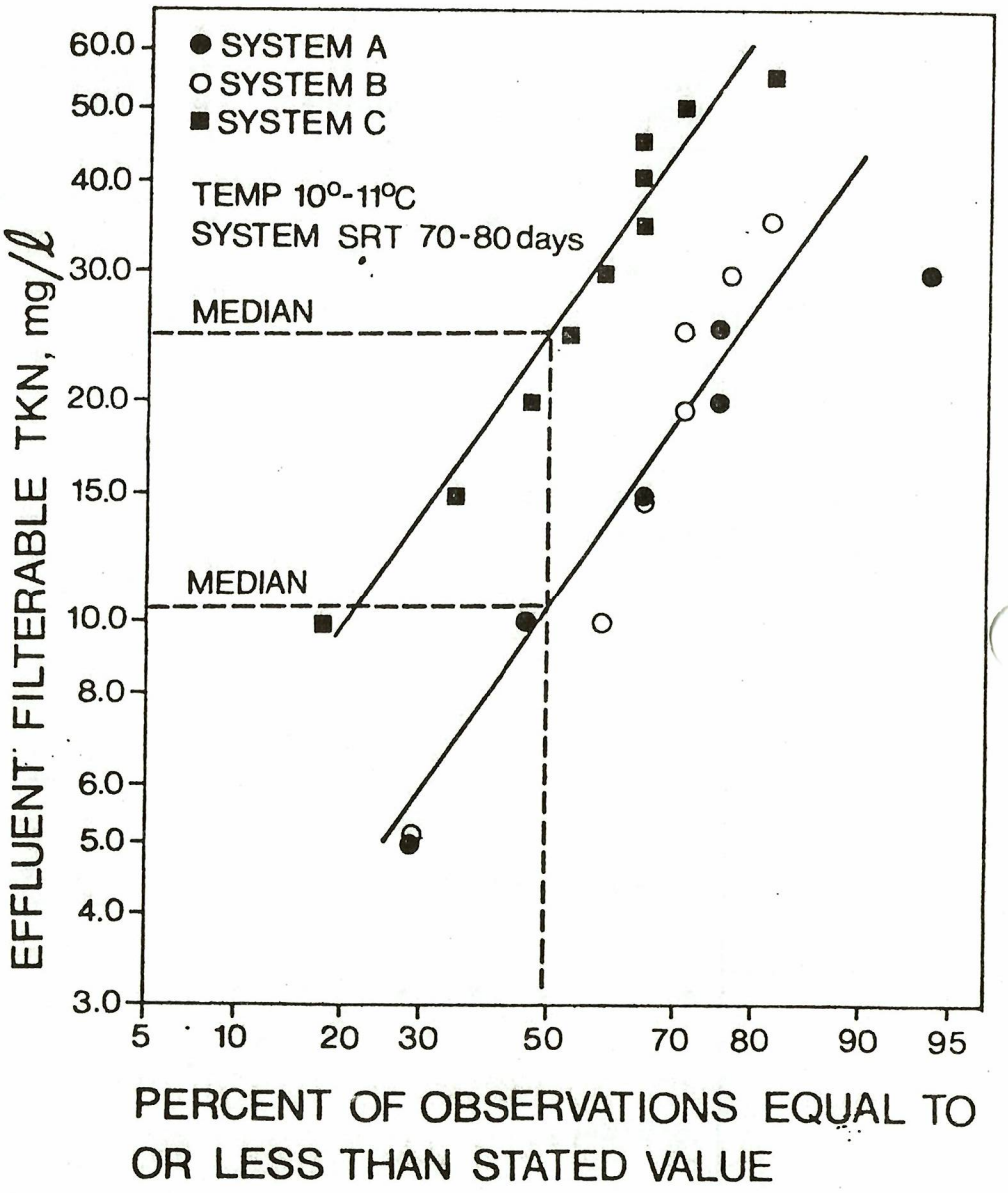


FIGURE 11. EFFLUENT FILTERABLE TKN RESULTS DURING LOW TEMPERATURE BENCH SCALE STUDIES

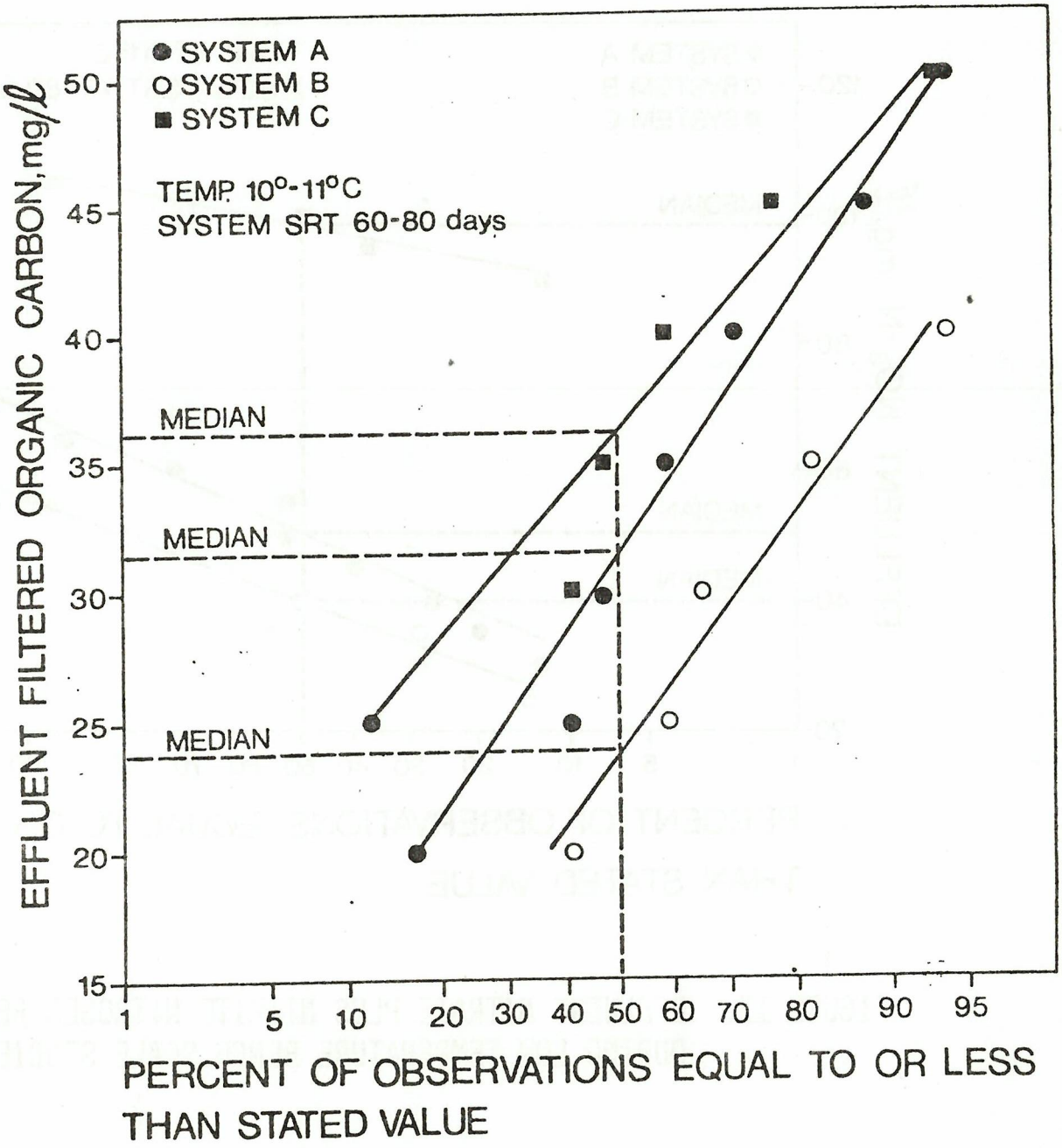


FIGURE 12. EFFLUENT FILTERABLE ORGANIC RESULTS DURING LOW TEMPERATURE BENCH SCALE STUDIES

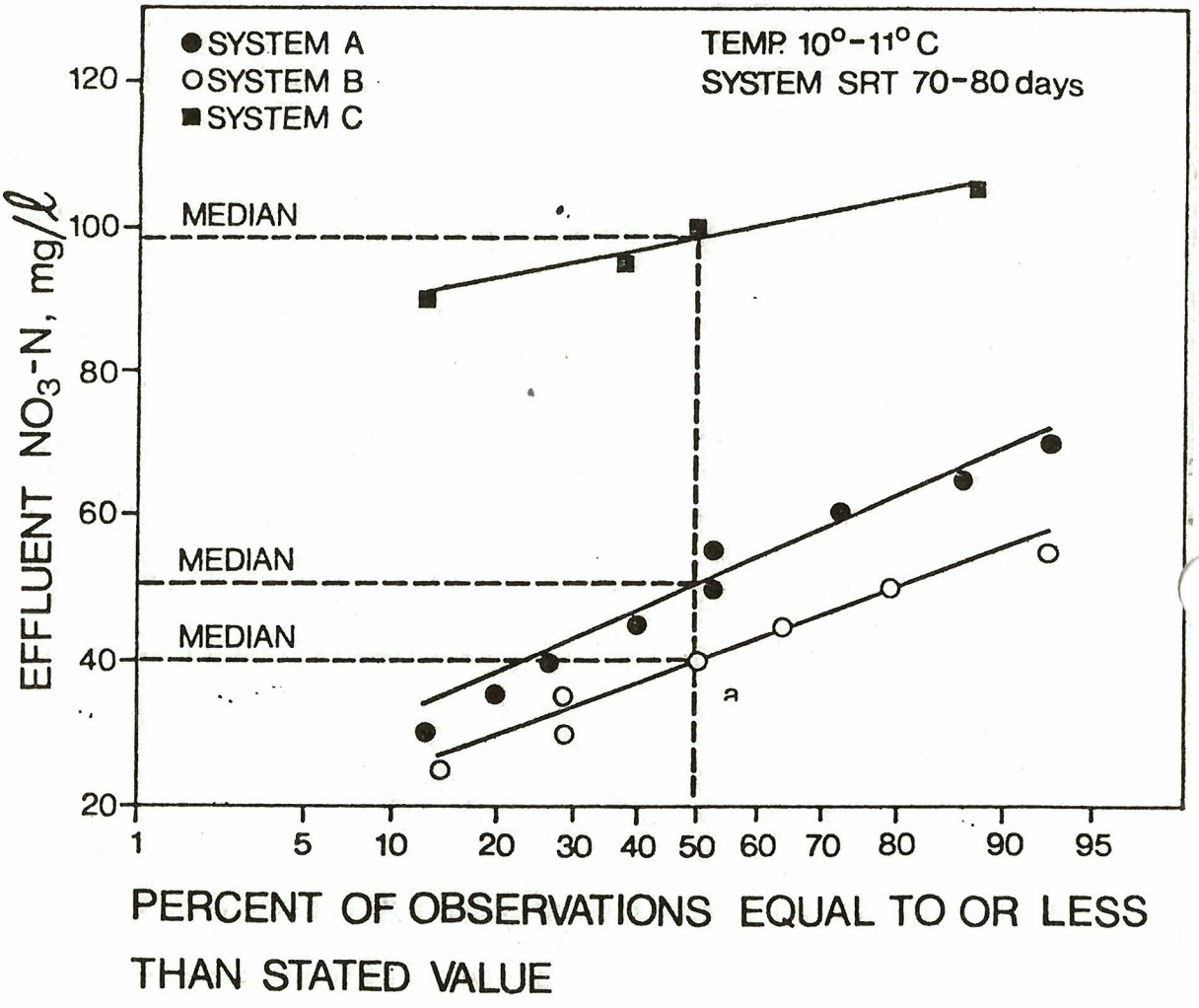


FIGURE 13. EFFLUENT NITRATE PLUS NITRITE-NITROGEN RESULTS DURING LOW TEMPERATURE BENCH SCALE STUDIES

(2) the nitrifier temperature sensitivity was greater than expected due to the effect on the organisms of exposure to a cyclic temperature change from 17° to 10°C in passing from the anoxic reactor to the aerobic reactor.

The phase 2 bench scale studies conducted following the 16 week operating period at 10°C were designed to assess the effect of the cyclic temperature change on the nitrification performance. During this period the systems were operated as two-stage single sludge systems, and the anoxic and aerobic reactors of systems B and C were operated at 17° and 10°C, respectively, as previously noted. The nitrification results from these studies were inconclusive. For a two to three week period following the initiation of the cyclic temperature conditions, the percent filterable TKN removal in all systems was above 90 percent. At this time a new batch of wastewater was obtained from Du Pont Maitland and after approximately one week, the percent filterable TKN removal in all bench scale systems had declined to less than 60 percent. Complete denitrifitation and a high degree of organic carbon removal were maintained during this period. Following these studies, time constraints prevented completion of any further experimental work.

Continuing studies by Du Pont of Canada Ltd.[11] have shown that the long-term solution to the winter nitrification problem will require increasing the minimum winter operating temperature of the aerobic basin using waste heat from the manufacturing process. During the winter of 1977-78, under conditions of good process control with SRT greater than 70 days, complete nitrification was not sustained. On the basis of this, Du Pont of Canada Ltd. has decided to proceed with a project to raise the aerobic basin temperature during the winter.

CONCLUSIONS

This study has identified a number of important factors for those designing and operating biological nitrification-denitrification systems.

Single sludge nitrification-denitrification systems can be designed to treat complex industrial wastes containing high concentrations of organic carbon, and organic and inorganic nitrogen. Design and operation of these facilities must be based on solids retention time and designers must think in terms of much higher SRT's than previously considered in order to achieve complete nitrification. Based on bench scale studies, the minimum aerobic SRT required to nitrify Du Pont wastewater was 25 to 30 days at 20°C, and 55 to 60 days at 10°C.

At aerobic basin temperatures of 10°C, autotrophic nitrifiers showed less tolerance for variations in influent composition and temperature than did heterotrophic organisms responsible for organic carbon oxidation and denitrification. It is likely that Du Pont of Canada Ltd. will continue to experience difficulty with winter nitrification until process changes are completed that will raise minimum winter aerobic basin temperatures[11].

Bench scale experiments completed in this study show that the performance of the Du Pont of Canada Ltd. process configuration was better than the two-stage separate sludge system. When powdered activated carbon was added to the Du Pont configuration the only measured change was a slight reduction in effluent filterable organic carbon.

Du Pont of Canada Ltd.'s experience with their full scale nitrification-denitrification process has demonstrated the viability of biological nitrogen removal as a low cost treatment method for complex wastes[11].

ACKNOWLEDGEMENTS

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REFERENCES

1. U.S.-EPA, "Process Design Manual for Nitrogen Control", Office of Technology Transfer, Washington, D.C. (1975).

2. Bridle, T.R., et al. "Start-up of a Full-Scale Nitrification-Denitrification Treatment Plant for Industrial Waste", 31st Purdue Industrial Waste Conference, Purdue University, Indiana (1976).

3. Levenspiel, O., Chemical Reaction Engineering, p 250, John Wiley and Sons Inc., New York (1967).

4. Sutton, P.M., et al. "Reliability of Nitrification Systems with Integrated Phosphorus Precipitation", Canada-Ontario Agreement Research Report No. 64 (1976)

5. Wild, H.E., et al. "Factors Affecting Nitrification Kinetics". J. Water Poll. Control Fed., 43, 1845 (1971).

6. Kiff, R.J., "The Ecology of Nitrification/Denitrification in Activated Sludge", J. Water Poll. Control Fed., 71, 475 (1972).

7. Wong-Chong, G.M., and Loehr, R.C., "Kinetics of Microbial Nitrification as Applied to the Treatment of Animal Waste - Ammonium-Nitrogen Oxidation", AIChE Symposium Series No. 145, Volume 71 (1975).

8. Barnard, J.L., "Cut P and N Without Chemicals", Water and Wastes Engineering, 11, 7, 33 (1974).

9. Sutton, P.M., et al. "Kinetic Studies of Single Sludge Nitrogen Removal Systems", IAWPR Conference on Advanced Treatment and Reclamation of Wastewater, Johannesburg, Republic of South Africa (1977).

10. Chow, V.T., Handbook of Applied Hydrology, p 8-13, McGraw-Hill, New York (1964).

11. Bridle, T.R., et al. "Operation of a Full-Scale Nitrification-Denitrification Industrial Waste Treatment Plant", 10th Mid-Atlantic Industrial Waste Conference, University of Delaware, Newark, Delaware (1978).

IMPACT OF CHEMICAL PHOSPHORUS REMOVAL ON AEROBIC AND ANAEROBIC DIGESTION OF SLUDGES

by

E.R. Hall

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NUTRIENT CONTROL TECHNOLOGY SEMINAR
Calgary, Alberta
February 7-8, 1980

Effects of chem precip of phosp

increased sludge mass & volume to be stabilized
increased organic loading
alkalinity & pH upsets
resolubilization of precipitated phosphorous
Toxicity of chemicals to methanogens & nitrifiers

INTRODUCTION

The addition of chemical coagulants to wastewater can be an effective and economical method of achieving improved suspended solids and phosphorus removal. However, lime, aluminum or iron salt addition to an existing biological treatment plant may lead to considerable changes in overall process performance. Of major interest is the effect of chemical addition upon sludge digestion systems which were originally designed for the stabilization of raw and biological sludges.

Chemical phosphorus removal causes major changes in sludge characteristics irregardless of the point of addition in the process. A number of these changes might be expected to significantly alter sludge digester performance. One such problem is attributable to the greater sludge masses and sludge volumes produced by chemical precipitation, which lead to increased solids and hydraulic loadings to the digester. The composition of a chemical sludge is difficult to predict. It will probably have a higher inorganic fraction, but because of the increased suspended solids removal, organic loadings to the digesters may also be greater. An existing digestion facility would typically reduce raw and activated sludge solids by 40 to 50%. Following the introduction of chemical phosphorus removal and an increase in inorganic solids, the fraction of solids that are biodegradable will decrease, perhaps to as low as 20 to 25%. With this relatively low solids reduction and in view of the larger sludge volumes requiring treatment, the utility of sludge digestion itself might be considered questionable.

Several other alterations in sludge quality may also lead to poor digester performance. Since anaeraobic digestion is particularly sensitive to reactor pH, it is important to consider the alkalinity and pH of the chemical sludge. Phosphorus removal with iron does not usually alter sludge pH drastically, but alum and lime precipitations are most effective at lower (<6.0) and higher (>9.0) pH's, respectively. An abrupt introduction of chemical sludges at these pH's into an operating digester may have inhibitory effects on the bacteria and may affect the stability of the chemical phosphate precipitate.

The phosphorus content of a chemical sludge is sufficiently high that even the release of a small fraction of the sludge phosphorus results in large quantities of soluble nutrient in digester supernatant. Since the supernatant is generally returned to the head of the treatment plant, phosphorus can begin to accumulate in the process, absorbing large quantities of additional chemicals. Therefore, consideration must be given to the potential for resolubilization of phosphorus during sludge digestion.

Finally, the toxicity of the chemicals to aerobic and anaerobic microorganisms should be assessed. Large quantities of Ca^{2+}, Fe^{3+} or Al^{3+} may significantly affect microbial activity, particularly the sensitive anaerobic methane-formers. The inclusion of metal toxicity as a separate point is made here to distinguish it from association with decreased biodegradability of the chemical sludge.

In attempting to examine some of these potential problems, this report uses case histories from a number of Water Pollution Control Plants (WPCP) in Ontario. These case histories are supplemented where possible by references to pertinent literature.

The unique opportunity to determine the effects of various chemical additions on the performance of full scale digesters in Ontario led to the initiation of a special survey by the Residue Management Section of the Wastewater Technology Centre, Burlington (Cohen and Hamoda, 1975). The survey involved the mailing of questionnaires to 195 WPCP's during June 1975, as well as follow-up site visits and interviews. Approximately half of the plants contacted were operated by the Ontario Ministry of the Environment (MOE) with the other half operated by various local, regional or private authorities.

Mail response was good with 132 WPCP superintendants returning completed questionnaires. Follow-up interviews increased the number of responses to 165. Of 139 WPCP's having mechanical treatment facilities, 25 (18%) were using aerobic digesters and 65 (47%) had heated anaerobic digesters.

Preliminary review of the information contained in the completed questionnaires revealed serious gaps in the data, particularly

from the non MOE plants. As indicated in the 1973 and 1974 MOE Operating Summaries, the number of WPC's operated by MOE increased by 2.6% between 1973 and 1974. During this period, the number of MOE plants using aerobic digestion increased by 80%, but the number of plants using anaerobic digesters declined by 6.4%. The impact of the energy crisis on this trend in future years remains to be seen.

Less than half of the plants using heated anaerobic digesters had information concerning sludge gas production. Virtually none of the plants measuring gas production monitored the methane or CO_2 content of the gas. During 1973, few plants were applying chemically precipitated sludges to their digesters, while nearly half the plants surveyed had initiated the chemical addition program by the end of 1974.

Of the 32 anaerobic digestion plants applying chemicals, iron salts were most frequently used (56%) followed by alum (35%) and lime (9%). A majority of the plants with two-stage digestion (61%) were applying chemicals on a routine basis compared to only 29% for plants with single stage digestion.

Aerobic digestion is becoming increasingly popular as an alternative to anaerobic digestion, particularly for smaller WPCP's. Operational data available from the 25 aerobic digestion plants surveyed is, however, woefully lacking. Only four plants surveyed had adequate information concerning sludge volumes, concentrations, and supernatant quality, and none of these plants were applying chemicals for phosphorus (P) removal. Of the eight aerobic digestion plants applying chemicals, five were using alum, and three iron salts.

Of the 90 WPCP's surveyed having digestion facilities, 11 were selected as case studies. A matrix was constructed (Table 1) of (a) digestion type (aerobic or anaerobic); (b) point of chemical addition (primary or secondary); (c) number of digestion stages (one or two); and (d) chemical used (alum, iron or lime). Alum and iron salts may be added to either the primary or secondary treatment units. Lime is invariably added to the primary process. Five different chemically mixed sludge types may therefore be produced for subsequent digestion, namely: (a) alum - primary sludge; (b) alum - secondary sludge;

TABLE 1. AEROBIC AND ANAEROBIC DIGESTION OF CHEMICALLY PRECIPITATED SLUDGES - TYPICAL PLANTS

	Anaerobic Digestion			Aerobic Digestion
Chemical Added	Primary Addition		Secondary Addition	Secondary Addition
	1 - Stage	2 - Stage	2 - Stage	
Alum	Pt. Edward	Barrie C.F.B. Petawawa	Kitchener	Bradford
Iron	[Pt. Edward]	Sarnia Chatham	N. Toronto Peterborough	-
Lime		Newmarket Midland	-	-

(c) iron - primary sludge; (d) iron - secondary sludge; and (e) lime - primary sludge. It should be noted that the alum and iron secondary sludges are usually returned to the primary clarifier to create a mixed chemical - primary + secondary sludge.

ANAEROBIC DIGESTION

A. Alum Sludges

Case Study 1 - Point Edward WPCP (0.2 MIGD)

Category - Alum - primary/anaerobic - single stage
Chemical dose - 300 mg/L alum (1972 - 1973)

Table 2 shows the dramatic deterioration in digester performance 90 days after initiation of alum phosphorus removal. Supernatant solids increased from 0.3% TS to 4.8%, doubling the recycled influent loading to the primary clarifier.

Because of this cumulative recycle effect, the final effluent suspended solids, BOD and total phosphorus concentrations also deteriorated to unacceptable levels.

Solutions Attempted (Sequentially)	Result
1) Lime addition (2 tons/6 months).	Negligible.
2) Reduce alum dosage to 200 mg/L.	Negligible.
3) Add anionic polymer (0.2 to 0.3 mg/L).	Slight improvement.
4) Reduce supernatant recycle to primary by increasing sludge truck haul from 1/week to 1/day.	Increased costs - poor effluent quality.
5) Substitute lime (120 mg/L) and $FeCl_3$ (20 mg/L as Fe^{3+}) for alum.	Dramatic recovery.

The quantitative results of solution (5) will be discussed separately under "Lime Sludges".

Case Study 2 - Barrie WPCP (3.0 MIGD)

Category - Alum - primary/anaerobic - two-stage
Chemical dose - 150 to 200 mg/L alum

Table 3 indicates the classic symptoms of anaerobic digester upset occurred with 200 mg/L alum including reduced gas production and increased volatile acid concentration. The reduced gas production resulted in a significant decline in digester temperature which was not immediately compensated for by adding an auxiliary fuel supply.

TABLE 2. PT. EDWARD WPCP - ANAEROBIC DIGESTION PERFORMANCE BEFORE AND AFTER ALUM ADDITION (300 mg/L)

Parameter	Without Alum			With Alum (After 90 days)		
	Raw Sludge	Digested Sludge	Super-natant	Raw Sludge	Digested Sludge	Super-natant
pH	6.5	7.2	7.0	5.8	5.8	5.8
% total solids	2.6	5.4	0.3	8.9	8.2	4.8
% Volatile fraction	76.0	63.0	61.0	69.2	68.2	67.7
Alkalinity (as $CaCO_3$) *		1420			930	
Volatile Acid (as CH_3COOH)*		88			903	
% CH_4 in gas		64			26	

* in mg/l

TABLE 3. BARRIE WPCP - ANAEROBIC DIGESTION PERFORMANCE BEFORE AND AFTER ALUM ADDITION (150 to 200 mg/L)

Parameter	Without Alum	With Alum	
		200 mg/l	150 mg/l
% T.S. Raw Sludge	4.54	2.76	4.35
Raw sludge volume $ft^3 \times 10^3$/day	14	32	16
Gas production $ft^3 \times 10^3$/day	29	20	37
Production ft^3/lb V.S. digested	16.3	9.0	17.0
Volatile acids (as CH_3COOH) mg/l	<1000	>2000	240-1200
Temp °F	90-95	80-85	90-95

Solutions Attempted (Sequentially)	Result
1) Reduce sludge loading from 0.10 to 0.08 lbs VS/ft^3/d.	Recovery within 30 days.
2) Add auxiliary fuel to raise temperature from 90 to 95°F.	Increased gas production created serious foaming condition. Rectified by mixing primary and secondary digested sludges.
3) Reduce alum dosage from 200 to 150 mg/L.	

Case Study 3 - CFB Petawawa WPCP (1.6 MIGD)

Reference: Shannon et al (1974)

Category - Alum - primary/anaerobic - two-stage

Chemical dose - 138 mg/L alum

The increase in solids produced and loaded to the digester coupled with the low alkalinity buffering capacity of the Petawawa sludge resulted in digester failure 7 days after initial alum addition (Figure 1). However, Figure 2 demonstrates that as the alum content of the digester sludge increases with time, the supernatant total and

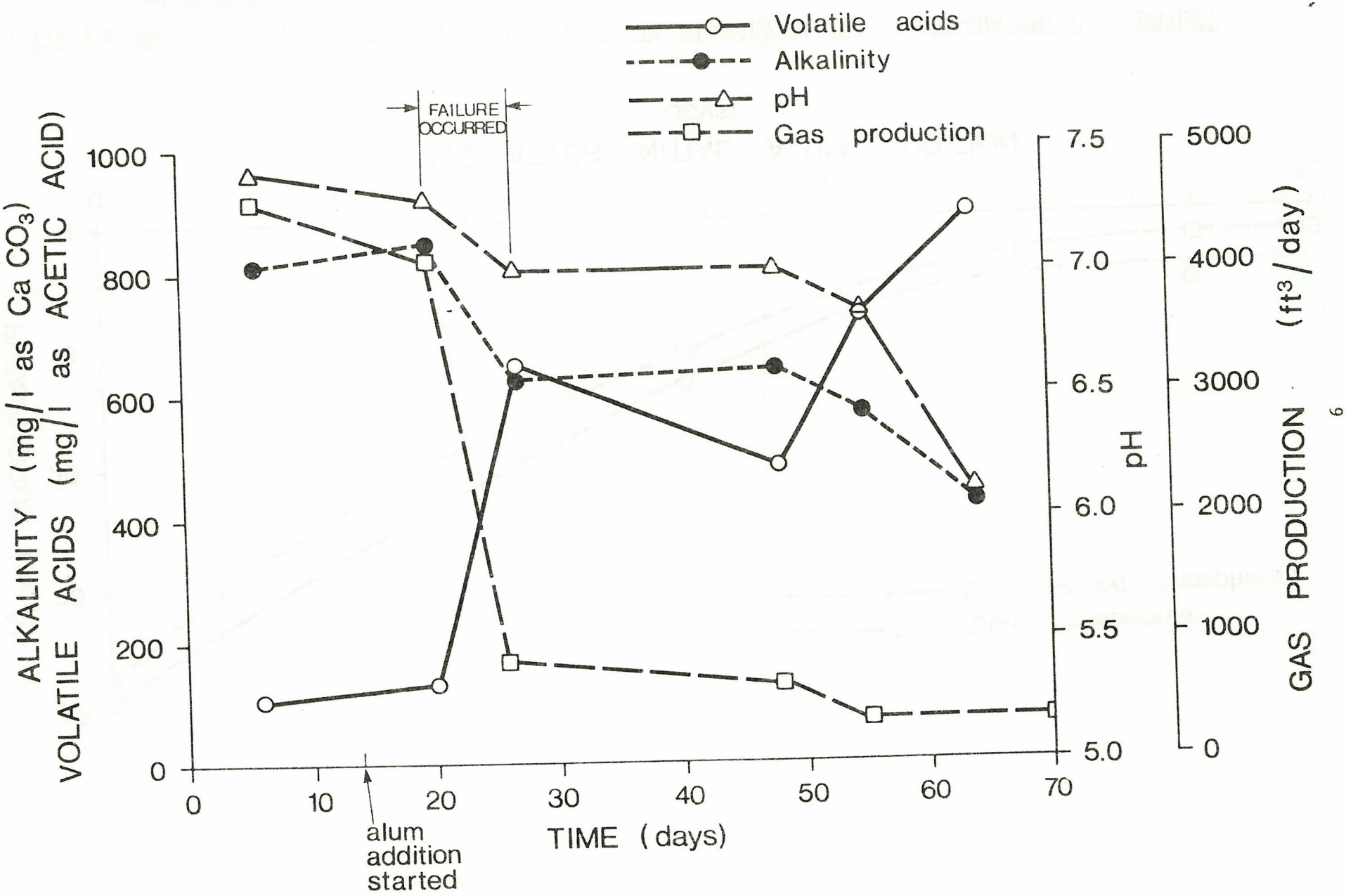


FIGURE 1 : ANAEROBIC DIGESTION OF ALUM PRIMARY SLUDGE (C.F.B. PETAWAWA, SHANNON ET AL, 1974)

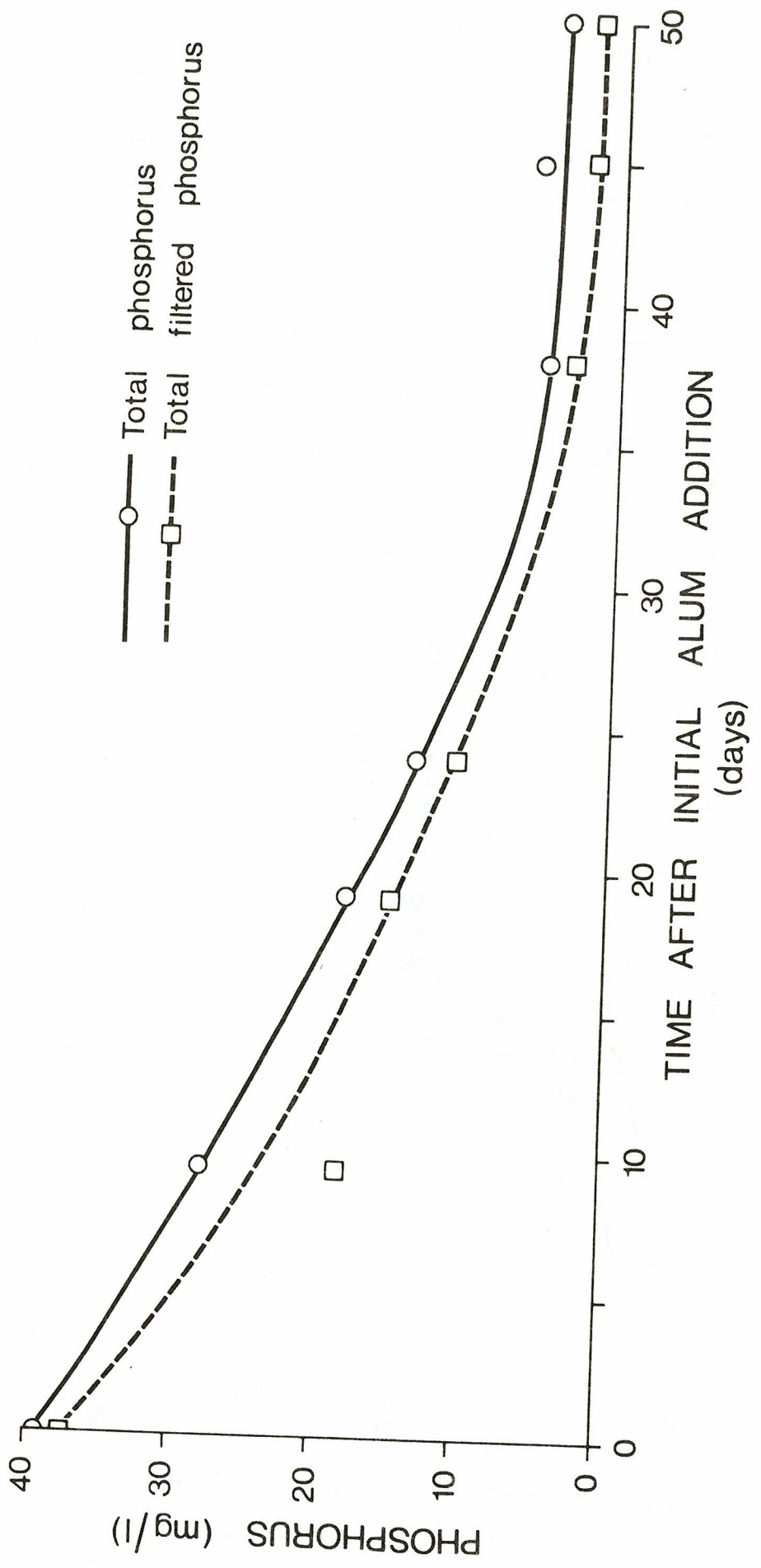


FIGURE 2: ANAEROBIC DIGESTION OF ALUM PRECIPITATED PRIMARY SLUDGE - SUPERNATANT P - CONTENT
(SHANNON ET AL, 1974)

filtered phosphorus concentration is reduced to 5 mg L^{-1} or less. The ability of the chemical sludge to retain this bound phosphorus in the solids fraction is apparently not affected by failure of the digestion process itself.

Solutions Attempted (Sequentially)	Result
1) Anionic polymer addition of 0.25 mg/L.	Negligible
2) Lime addition.	Negligible
3) Dump digester and restart with incremental alum dosages (30 mg/L) stepwise; 2 to 3-week intervals at each increasing alum dosage for microbial acclimation.	

Auxiliary fuel supply helped maintain a constant temperature (95°F) throughout the 70-day field trial. Soluble aluminum ions were measured in the digester supernatant before and after digester failure occurred. The slight increase from 0.5 to 2.0 mg/L does not indicate that aluminum toxicity per se was the major cause of failure. The sudden increase in voaltile solids loading from 0.028 to 0.051 lb VS/ft^3/d is therefore considered to be the major stress factor.

Case Study 4 - Kitchener WPCP (14 MIGD)

Category - Alum - primary/anaerobic - two-stage
Chemical dose - 80 to 120 mg/L alum

Table 4 indicates that the most cost effective method of achieving a final effluent total-P concentration of less than 1.0 mg/L was to add 105 mg/L of alum to the mixed liquor entering the secondary clarifier.

On the basis of these results, full scale alum addition was inaugurated in late 1972. The initial alum dose was 120 mg/L, but by early 1974 this was reduced to 80 mg/L. A foaming problem in the mixed liquor tanks and primary digester during the first two weeks of alum

application was solved by changing the return sludge rate to increase the mixed liquor volatile suspended solids (MLVSS) concentration. The change in digestion performance before and after alum addition is summarized in Table 5.

TABLE 4. KITCHENER WPCP - JAR TEST OPTIMAL DOSAGES AND COSTS FOR P-REMOVAL WITH ALUM, IRON, AND LIME CHEMICALS

Chemical	Primary Addition		Secondary Addition	
	mg/l	$/M.I.G.	mg/l	$/M.I.G.
Alum	300	90	105	31
Ferric Chloride	52	78	22	33
Lime	500	37	-	-

With the exception of gas production quantity and quality, for which no data are available, the application of approximately 100 mg/L alum had a beneficial effect on digester performance, including solids concentration, volatile fraction reduction, and supernatant quality. The increase in alum secondary sludge weight of 26.7% was offset to some extent by the increase in solids concentration which resulted in an 11% decline in digested sludge volumes requiring ultimate disposal.

The Kitchener WPCP operator interviewed indicated that the number of operational problems at the plant have increased since alum addition, particularly with foaming in the overloaded primary digester. The recirculation pumps have difficulty in handling the foaming sludge, leading to occasional non-homogeneous mixing and heat distribution problems in the digester. Despite these problems, however, the Kitchener plant is the most successful case of full scale alum sludge anaerobic digestion in Ontario.

Some of the factors that may account for the relatively successful application of alum sludge at Kitchener, compared to the Point Edward, Barrie and CFB Petawawa case studies, are: (a) higher

TABLE 5. KITCHENER WPCP - ANAEROBIC DIGESTION PERFORMANCE BEFORE AND AFTER ALUM ADDITION (80 to 120 mg/L)

Parameter	Without Alum (55 days)	With Alum (55 days)	% Change
Waste sludge (WAS) volume M.I.G.	9.47	9.15	-3.5
WAS weight (lbs x 10^3)	609	837	+37.4
WAS Vol. Ratio (IG/MIG)	12,616	11,233	-11.0
WAS Wt. Ratio (lbs/MIG)	811	1,028	+26.7
Raw sludge % total solids	5.0	5.3	+6.0
Raw sludge % volatile fraction	66.0	61.0	-7.6
Digested sludge % total solids	3.5	4.3	+22.8
Digested Sludge % volatile fraction	56.0	49.0	-12.5
Overall V.S. reduction %	34	39	+14.7
Supernatant Quality % T.S.	1.4	0.1	-93.0

alkalinity - buffered sludge; (b) low alum dose required for P-removal (80 mg/L); and (c) application to secondary rather than primary unit process.

Literature Review

Similar to the experience in Ontario, Zenz and Pivnicka (1969) reported that alum coagulation resulted in larger sludge volumes requiring digestion, but that waste activated sludge from the alum dosed aeration tank did not impair the digestion process. It was also noted that although a digester receiving alum sludge had an expected higher phosphorus content, the bound phosphorus removed by chemical-biological treatment was not released during anaerobic digestion. A similar result was also reported by Barth and Ettinger (1967).

A laboratory scale study of the digestion of aluminum hydroxide sludges (Hsu and Pipes, 1973) demonstrated that when an Al concentration was greater than 200 mg/L, anaerobic digestion as monitored by gas production was depressed. Since the digester pH remained between 7.0 to 7.6, the effect of aluminum hydroxide was concluded to be on the volatile acid formation step.

Gossett et al (1978) also found that a full scale digester was inhibited when alum was added at 200 to 400 mg/L. Under these conditions gas production and volatile solids and COD removals were reduced (Figure 3), along with sludge alkalinity (Table 6). It was suspected that lowered alkalinity resulted from reduced degradation of organic nitrogen compounds. In anaerobic digestion, alkalinity is formed from ammonia released from organic compounds:

$$NH_3 + CO_2 + H_2O \rightleftharpoons HCO_3^- + NH_4 \qquad (1)$$

As shown in Table 7, organic nitrogen decomposition was only one-half as great for alum as for control sludges. The effect of alum addition was concluded to be a depression of the kinetic rate of organic degradation, in agreement with the observations of Hsu and Pipes (1973).

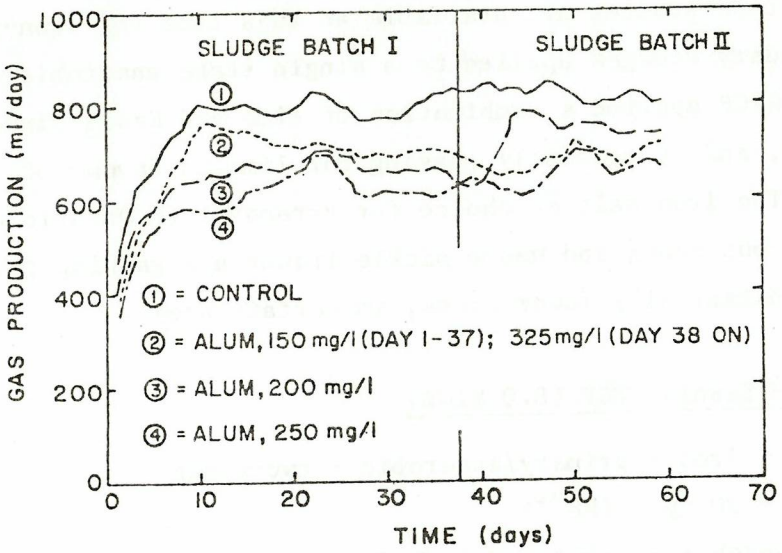


FIGURE 3: DAILY GAS PRODUCTIONS FROM ALUM SLUDGES (GOSSETT ET AL, 1978)

TABLE 6. DIGESTER ALKALINITY, VOLATILE ACIDS, AND pH FOR SLUDGES DERIVED FROM ALUM ADDITION TO WASTEWATER*

Parameter	Control Value	Alum 200 mg/L	Alum 250 mg/L	Alum 325 mg/L	Alum 400 mg/L
Alkalinity (mg/L as $CaCO_3$)	2450	2350	1800	1950	1050
Volatile Acids (mg/L as acetic acid)	100	100	125	100	425
pH	6.9	6.8	6.7	6.7	6.0

* Gossett et al (1978)

TABLE 7. NITROGEN TRANSFORMATIONS DURING DIGESTION*

Sludge	SRT (days)	Organic-N Reduction (%)
Control	15	31
Alum, 325 mg/L	15	14
Alum, 400 mg/L	15	12
Control	30	35
Alum, 325 mg/L	30	16

* Gossett et al (1978).

B. Iron Sludges

No case studies are available at this time for iron-primary or iron-secondary sludges applied to a single stage anaerobic digester. Point Edward WPCP applied a combination of lime and $FeCl_3$ since December 1973, and is presently phasing out lime in favour of iron salts only. The iron salt of choice for P-removal in Ontario was initially $FeCl_3$, but $FeSO_4$ and waste pickle liquor are gaining favour because of substantially lower costs, in certain areas.

Case Study 1 - Sarnia WPCP (8.0 MIGD)

Category - Iron - primary/anaerobic - two-stage

Chemical dose - 20 mg/L (Fe^{3+})

Although comparative data before and after iron addition are not available for this plant, Table 8 indicates that there have been no noticeably adverse effects on anaerobic digestion performance since chemical addition was initiated.

TABLE 8. SARNIA WPCP - ANAEROBIC DIGESTION PERFORMANCE AFTER IRON ADDITION (20 mg/L Fe^{3+})

Parameter	Raw Sludge	Digested Sludge	Supernatant	Remarks
% total solids	4.4	16	0.17	
% volatile fraction	54.5	39.4	-	45% V.S. reduction
pH	6.9	7.0	7.0	
Total-P (mg/l)	1120	3000	30	
Soluble-P (mg/l)	1.0	1.9	0.5	
% CH_4 in gas		68		

Gas production is reported to have increased from 60 000 ft^3/d before iron addition to 72 000 ft^3/d after iron addition. This increase reflects the increased solids loading to the digester which even after iron addition amounted to only 0.04 lbs VS/ft^3/d.

The increase in gas production after iron addition enabled 100% burning of methane for heating and fuel purposes, whereas 25% auxiliary fuel was required by the WPCP before iron addition.

The major problem encountered at Sarnia since iron addition commenced involved generaton of a high volatile acid condition (1400 mg/L) resulting from poor mixing of the very heavy (16% TS) sludge. The gas recirculation equipment was not capable of handling this increase in solids concentration causing digested sludge to settle below the inlet to the sludge recirculation pumps.

Solutions Attempted (Concurrently)	Result
1) Withdraw sludge from bottom port.	Satisfactory.
2) Install larger sludge recirculation pumps.	

The very low phosphorus concentrations found in the supernatant recycle indicate that once precipitated P is bound to iron, anaerobic digestion up to 45 days did not significantly resolubilize the phosphorus.

Case Study 2 - Chatham WPCP (4.5 MIGD)

Category - Iron - primary/anaerobic - two-stage

Chemical dose - $FeSO_4$ - 16 mg/L as Fe (since January 1974)

Table 9 shows that since iron was added to the primary treatment process, the total solids concentration of the raw sludge declined by 26.3%, while the digested sludge solids concentration declined by 9.8%. No significant changes in digester pH, alkalinity or volatile acids were noted after iron addition. In July 1974, a waste iron salt solution was substituted for commercial grade $FeSO_4$. No changes in digester performance were noted after this substitution.

Only one serious digester upset has occurred at this plant since iron addition was started. The volatile acids on this occasion rose to 1800 mg/L. Plant personnel recognized the problem as being due to a temporary overloading condition. No lime was added.

TABLE 9. CHATHAM WPCP - ANAEROBIC DIGESTION PERFORMANCE BEFORE AND AFTER IRON ADDITION (16 mg/L Fe^{2+})

Parameter	Without Iron (1973)	With Iron (1974)
Raw Sludge % total solids	5.7	4.2
Raw Sludge % volatile fraction	52	50
Digested Sludge % T.S.	5.1	4.6
Digested Sludge % volatile fraction	40	39
Supernatant % T.S.	1.8	1.4
Sludge haul yd^3/month	3181	3155
Gas production $ft^3 \times 10^3$/month	41	45

Solution Attempted — Result

Reduce loading. — Full recovery in 12 days.

Centrifugation tests were conducted at Chatham on digested sludged with iron addition to the primary process. No significant differences were observed in dewaterability that could be related to iron addition.

Case Study 3 - North Toronto WPCP (7.5 MIGD)

Category - Iron - secondary/anaerobic - two-stage

Chemical dose - 12 mg/L as Fe^{3+}

Unlike the situation at Kitchener where ferric chloride demand for primary addition was 2.4 times the demand for secondary

addition, no differences in this demand were noted at North Toronto. Indeed, $FeCl_3$ was first applied to the primary process and then shifted to the secondary process at the same dosage with no adverse effects on either P-removal or digester performance. Before iron addition, total solids concentrations of both raw and digested sludges averaged 6.0%. After iron addition, the raw sludge solids concentration declined to 3% TS. The iron improved sludge settleability to the extent that the average percent TS (after supernating) of the digested sludge increased from 6.0 to 9.0%. Thus the ratio of digested sludge % TS to raw sludge % TS increased threefold after iron addition. No adverse effects were reported for subsequent vacuum filtration of the iron digested sludge compared to vacuum filter performance with non iron digested sludge. Although sludge gas quantity and quality data are not available, a report by Van Fleet et al (1974) indicates a slight reduction in gas production occurred following introduction of $FeCl_3$ for P-removal.

Case Study 4 - Peterborough WPCP (9.0 MIGD)

Category - Iron - secondary/anaerobic - two-stage

Chemical dose - 9 to 12 mg/L as Fe^{3+}

Table 10 summarizes the impact of ferric chloride addition for P-removal on the anaerobic digestion process.

With the exception of improved supernatant quality and an increase in gas production following introduction of iron, no significant changes in digestion performance were noted. No digester upsets were recorded for the period (February 1974 - June 1975).

In summary, all four case studies involving addition of iron salts between 9 to 16 mg/L as Fe showed no significant deleterious effects and in some cases an improvement in anaerobic digestion performance.

TABLE 10. PETERBOROUGH WPCP - ANAEROBIC DIGESTION PERFORMANCE BEFORE AND AFTER IRON ADDITION (9 to 12 mg/L as Fe^{3+})

Parameter	Before Iron	After Iron
Raw Sludge % total solids	3.7	3.8
Raw Sludge % volatile fraction	75.5	67.7
Raw Sludge Volume MIG/yr	9.85	12.12
Digested Sludge % T.S.	2.4	2.5
Digested Sludge % volatile fraction	61.0	53.3
Volatile Acids mg/l as CH_3COOH	600	680
Supernatant T.S.S. (mg/l)	100	50
Gas Production $ft^3 \times 10^3$/day	68	75
Gas Quality % CH_4	65-70	65-70

Literature Review

Studies with ferrous chloride sludges by Malhotra et al (1971) have shown that iron sludges are quite compatible with anaerobic treatment. Volatile solids reduction, pH and volatile solids and alkalinity levels all indicated similar digester behaviour with chemical and non-chemical sludges. However, contrary to the experience at the Sarnia WPCP, significant quantities of phosphorus were released from digester solids during treatment. This resolubilization was greater for sludges containing higher concentrations of iron precipitated phosphorus.

Sludge digestion rate with ferric chloride sludge was found to be reduced considerably in comparison to a control system by Gossett et al (1978). The effect on biodegradation was thought to result from a "cage" effect in which the chemical floc forms a chemical or physical barrier to enzymatic hydrolysis of the volatile solids.

C. Lime Sludges

Case Study 1 - Point Edward WPCP (0.2 MIGD)

Category - Lime + ferric primary/anaerobic - single stage

Chemical dose - 120 mg/L lime + 20 mg/L $FeCl_3$

Table 11 compares the effect of lime + ferric addition on the anaerobic digestion process with the previous alum addition.

The substitution of lime and ferric chloride for alum had a dramatic result within 30 days. Supernatant recycle load dropped fivefold and methane conversion increased nearly threefold from 26% of total gas composition to 69% compared to the pre-chemical addition gas composition of 64% CH_4.

Some lime scale build-up problems were encountered after conversion of the system to lime + ferric. When the pH of the feed sludge was dropped below 8.5 the scale build-up problem could be reduced to an acceptable level. The pH adjustment was accomplished by gradually reducing the lime dosage while concurrently increasing the ferric chloride dosage. The problems involved in controlling a double chemical P-removal system led to the decision to apply ferric chloride only during 1975.

Case Study 2 - Midland WPCP (1.7 MIGD)

Category - Lime - primary/anaerobic - two-stage

Chemical dose - 125 mg/L lime

Numerous problems have plagued the Midland WPCP since inauguration of the lime addition program in January 1974. Lime build-up in the lines in contact with lime-primary sludge has seriously reduced throughput capacity. Sludge volumes have increased fivefold

TABLE 11. PT. EDWARD WPCP - ANAEROBIC DIGESTION PERFORMANCE AFTER ALUM ADDITION VERSUS DIGESTION AFTER LIME AND FERRIC ADDITION

Parameter	With Alum (90 days)			With Lime and Ferric (90 days)		
	Raw Sludge	Digested Sludge	Super-natant	Raw Sludge	Digested Sludge	Super-natant
pH	5.8	5.8	5.8	8.0	7.2	7.3
% T.S.	8.9	8.2	4.8	5.8	6.0	0.9
% volatile fraction	69.2	68.2	67.7	54.0	45.2	49.8
Alkalinity (as $CaCO_3$) mg/l	-	930	-	-	2600	-
Volatile Acids (as CH_3COOH) mg/l	-	903	-	-	421	-
% CH_4 in gas	-	26	-	-	69	-

increasing the cost of haulage and disposal. Foaming has occurred in both final effluent and primary digester. Supernatant quality has deteriorated (Table 12). The consistency of the lime feed slurry concentration (30% w/w) has been erratic. Finally, the original objective of satisfactory P-removal has not been achieved (only 50% removal to 5.7 mg/L total P at 125 mg/L lime). A complicating factor in the analysis of this problem is the possibility that metal plating industrial waste having high (no analytical data available) nickel concentrations may be exacerbating this condition.

TABLE 12. MIDLAND WPCP - ANAEROBIC DIGESTION PERFORMANCE BEFORE AND AFTER LIME ADDITION (125 mg/L)

Parameter	Before Lime (1972)	After Lime (1974)
Raw Sludge % T.S.	11.9	8.4
Raw Sludge % volatile fraction	46	30
Digested Sludge % T.S.	8.1	8.4
Digested Sludge % volatile fraction	34	24
Supernatant % T.S.	2.0	2.8
Sludge Haul yd^3/month	161	826

Solution Attempted	Result
Dump digesters once/year and restart.	Not satisfactory.

The Midland WPCP operator interviewed commented that he must pump inordinate volumes of lime-primary sludge to prevent sludge blanket build-up because thick lime sludge tends to stick to the bottom of the primary clarifier and resist removal by scraping.

Case Study 3 - Newmarket WPCP (2.0 MIGD)

Category - Lime - primary/anaerobic - two-stage

Chemical dose - 150 to 200 mg/L $Ca(OH)_2$

This plant was the first in Ontario (March 1971) to apply lime for P-removal. Digester operation at the time of lime addition start up was considered "normal" with good gas production and composition. According to Black (1973), lime dosage during the first 90 days was very erratic ranging between 0 to 500 mg/L. The pH of the sludge increased beyond the capacity of CO_2 production to neutralize, and by day 105, the digester had to be dumped.

Solution Attempted (Concurrently)	Result
1) Modification of raw sludge pumping schedule.	Gas production (76% CH_4) increased to provide 50% of boiler fuel requirements.
2) New pump installed to allow recycle of lime sludge back to rapid mix tank.	
3) Lime sludge blanket maintained at a depth of 1.5 to 2.0 ft.	
4) Digesters refilled and heat exchangers turned on February 1972.	

Despite the improvement in digestion performance noted after the February 1972 restart, operational problems still exist. These include: (a) constancy of lime feed dosage; (b) bulking in aerator requiring intermittent lime shut-off; (c) extremely concentrated ($\simeq$30%) sludge bottom layer causing sludge withdrawal problems; (d) very great variability in hauled sludge solids concentration (3 to 30%); (e) grit accumulation on bottom of digester interfering with mixing efficiency and reducing actual digester capacity; and (f) $CaCO_3$ deposits on all piping and channeling in contact with lime sludge. After hydrated lime was substituted for carbide waste lime, this problem diminished somewhat but is still bothersome. Flushing once a week and a major clean up once per month are required to keep the scale build-up problem under control.

Pilot centrifugation with a 200 mg/L dosed lime sludge and 1 lb/ton cationic polymer produced a good cake (27 to 34% TS) with high recovery rate (95 to 98%) and good centrate quality ($<$700 mg/L TSS). Phosphorus resolubilization after anaerobic digestion of the lime sludge was minimal.

Literature Review

As found at the Newmarket WPCP, a major problem associated with the anaerobic digestion of lime sludges is the high pH. Tofflemire and Hetling (1973) found that sludge pH will decrease considerably during the first 24 hours of existence. Van Fleet et al (1974) also indicated that the pH of lime sludge was often 10 to 10.5. By increasing the residence time of the solids in the clarifier to 1 to 1.5 days, sludge pH dropped to 9.5 and digester pH could then be maintained in the range 7.2 to 7.4. To avoid pH shocks to anaerobic digesters, Black (1976) suggested that lime addition should be increased stepwise over a three to four week period until the optimum level is reached.

AEROBIC DIGESTION

The fact that the aerobic stabilization process is much less susceptible to "upsets" than is the anaerobic process is well documented in the literature (Vesilind, 1974). Aerobic volatile suspended solids reduction can be predicted on the basis of a sludge retention time-temperature factor (Cohen, 1974).

Unfortunately, operating data for full scale aerobic digestion are scarce and for chemically precipitated sludges virtually non-existent.

The only case study available to date (Black, 1975) will be supplemented by information gleaned from the literature.

1. Alum and Iron Sludges

Case Study 1 - Bradford WPCP (0.9 MIGD)

Category - Alum - secondary/aerobic

Chemical dose - 100 mg/L alum

Table 13 compares aerobic digestion performance with and without alum addition.

TABLE 13. BRADFORD WPCP - AEROBIC DIGESTION PERFORMANCE WITH AND WITHOUT ALUM ADDITION

Parameter	Without Alum	With Alum
Raw Sludge % T.S.	1.3	0.6
Digested Sludge % T.S.	1.8	2.7
Thickening Ratio ($\frac{\text{Digested}}{\text{Raw}}$)	1.38	4.50
Supernatant T.S.S. mg/l	40-100	30-80
" BOD "	15-20	12-15
" Total P "	35	3-10
" Soluble P "	-	2.7
% V.S. Reduction	27	55

The very long detention time (138 days SRT), coupled with the alum addition of 200 mg/L, have created conditions that allow for excellent gravity thickening in the continuously fed clarifier-thickener hydraulically loaded at 70 gal/ft^2/d. Digester supernatant quality was good, particularly with regard to total and soluble phosphorus concentrations after alum addition.

Batch aerobic digestion studies comparing alum secondary sludge (200 mg/L) and iron secondary (10 mg/L Fe^{3+}) with a non-chemical control sludge (Ganczarczyk and Hamoda, 1973) indicate that the kinetic digestion rate constant for oxidizable VSS reduction was 0.067 for alum, 0.048 for iron and 0.037 for the control (Figure 4). When the alum dosage was increased to 550 mg/L, however, the alum sludge digestion performance was poorer than iron sludge (30 mg/L as Fe^{3+}), and the control. Eikum et al (1974) compared the VSS reduction rate of an alum primary sludge and a non-chemical control (Figure 5). He found that at short detention times (five days) the alum sludge digestion rate was greater than the control, particuarly at higher temperatures (25°C). With longer detention times (25 days), however, the situation was reversed with VSS reduction of 45% for the alum sludge versus 60% VSS reduction for the control at 25°C.

Ganczarczyk and Hamoda (1973) also studied the effect of 200 mg/L alum and 10 mg/L Fe^{3+} on the degree of nitrification as measured by NO_3-N concentration in the digester supernatant. Figure 6 shows that nitrification was not adversely affected by the aluminum or the iron in the sludge.

2. Lime Sludges

Hamoda and Ganczarczyk (1977) investigated the aerobic decomposition of lime precipitated primary sludge (Figure 7). A lime dosage of 200 mg/L resulted in VSS reduction superior to that of the control. Higher lime dosages, however, had an adverse effect on oxidizable VSS reduction. The difference was attributed to an improved precipitation of easily degradable organics under the lower lime conditions.

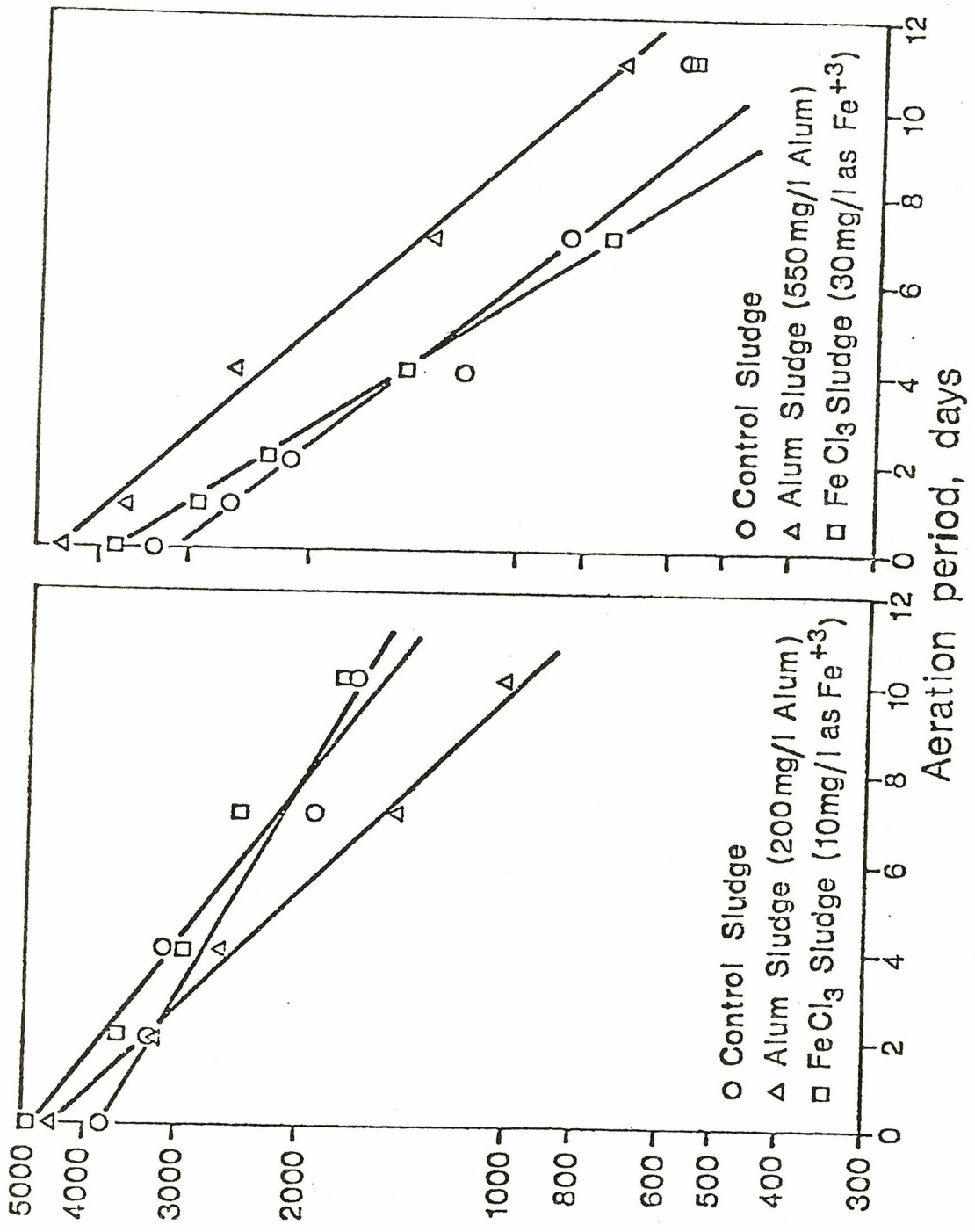


FIGURE 4: AEROBIC DIGESTION OF ALUM AND FERRIC CHLORIDE SLUDGES (GANCZARCZYK AND HAMODA, 1973)

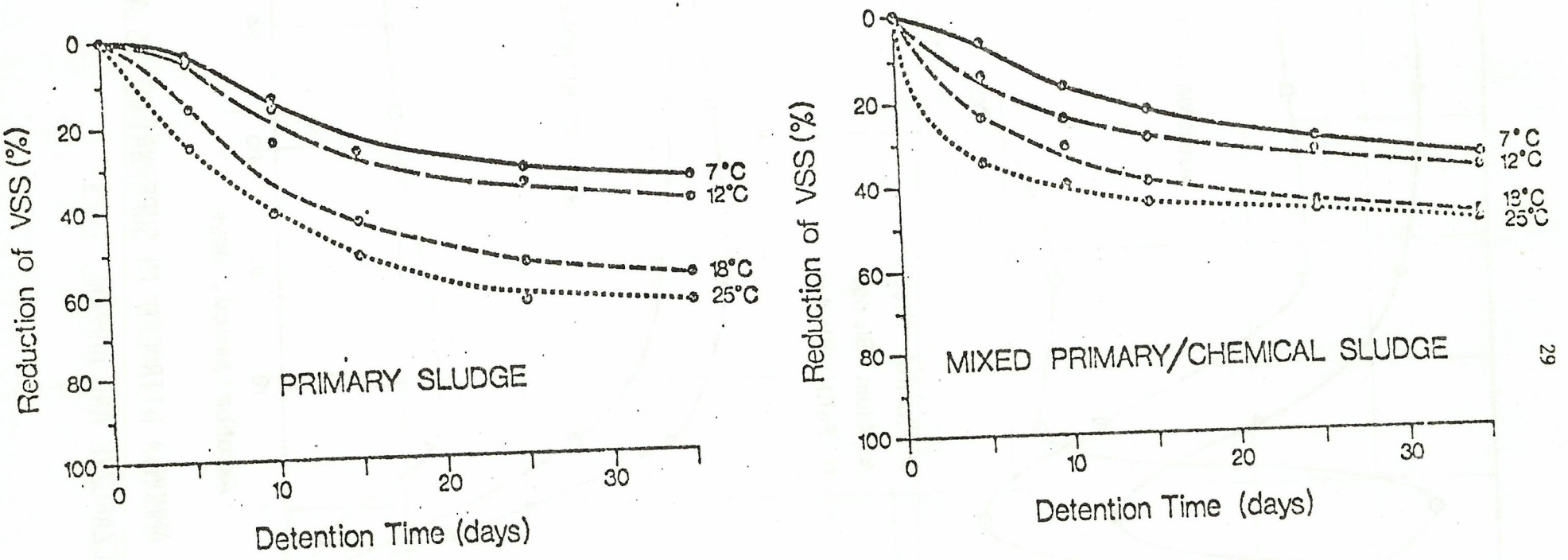


FIGURE 5: AEROBIC DIGESTION OF ALUM PRIMARY SLUDGE - % VSS REDUCTION (EIKUM ET AL, 1974)

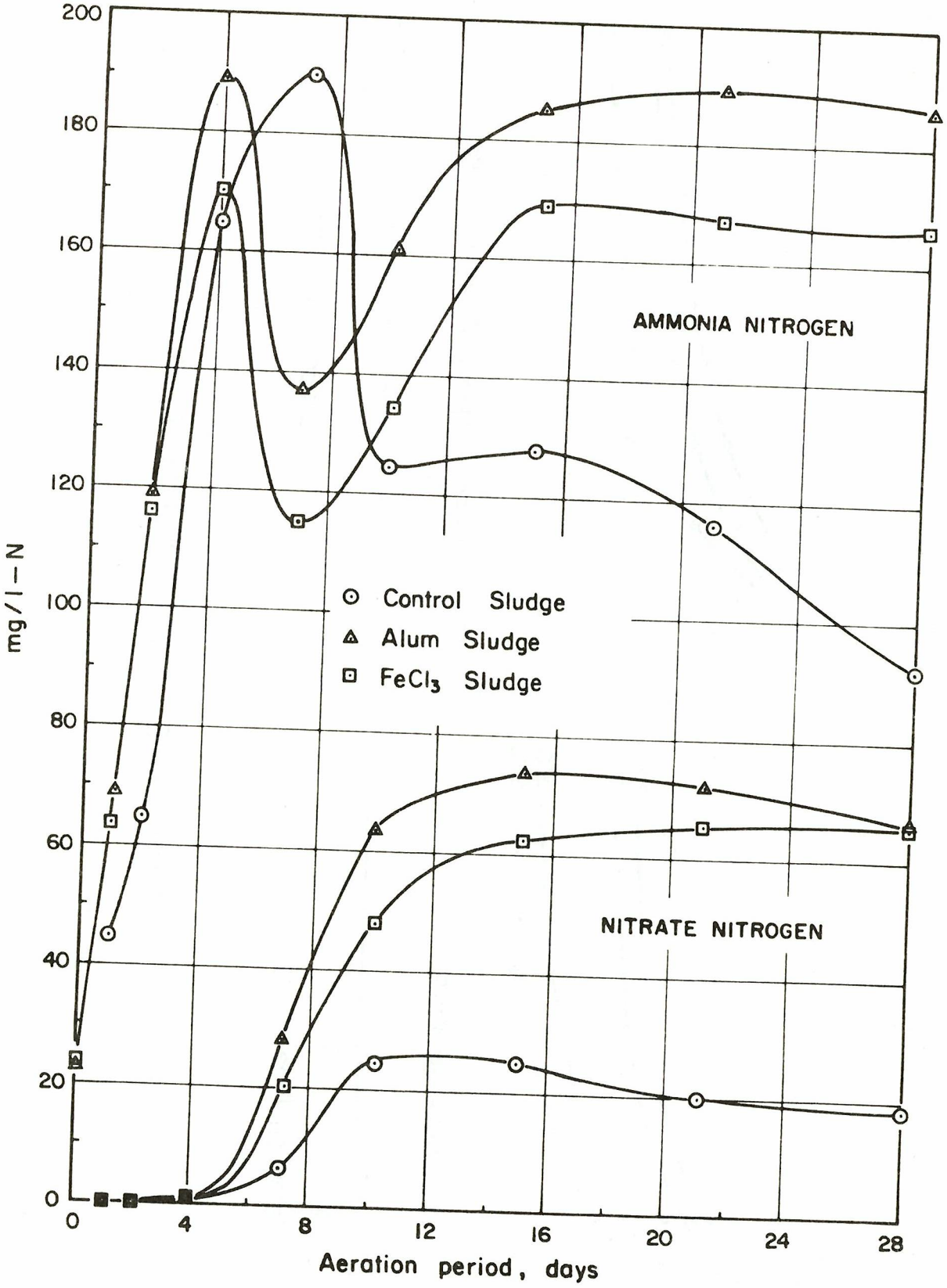
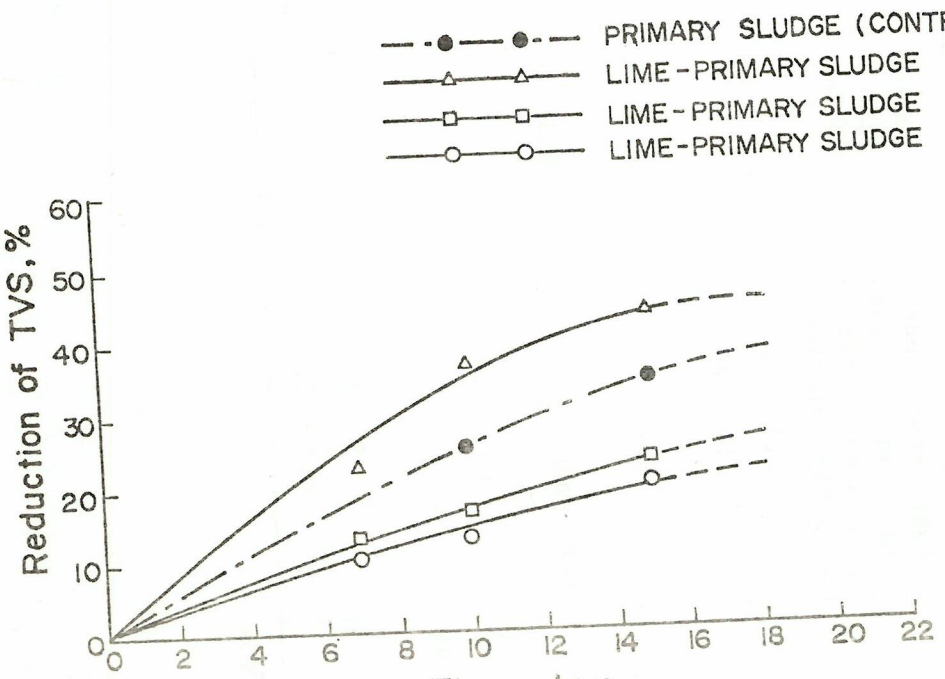
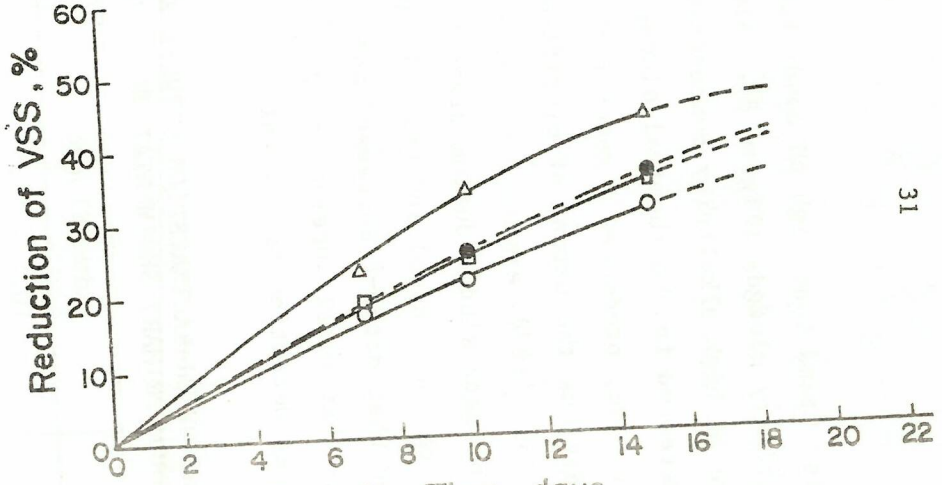


FIGURE 6: NITRATE AND AMMONIA NITROGEN IN SUPERNATANT VS AERATION PERIOD (GANCZARCZYK AND HAMODA, 1973)



REDUCTION OF TOTAL VOLATILE SOLIDS (TVS)
VS. DETENTION TIME



REDUCTION OF VOLATILE SUSPENDED SOLIDS (VSS.
VS. DETENTION TIME

FIGURE 7: REDUCTION DURING AEROBIC DIGESTION OF LIME - PRIMARY SLUDGE (HAMODA AND GANCZARCZYK, 1977)

It was also found that high pH conditions persisted during digestion of lime primary sludges (Figure 8). Although the elevated pH's were indicative of high alkalinity concentrations, very little nitrification was detected in the chemical sludge digesters. The control sludge, without lime, producted a reactor pH range of 6.0 to 7.2 and under these conditions the extent of nitrification was appreciable.

Huang and Tsai (1978) examined the potential for phosphorus release from a lime primary sludge during digestion. The physical and chemical composition of the control and test sludges are indicated in Table 14. They found that ortho-P increased about 1 mg/L in all digester supernatants, but that total supernatant phosphorus was reduced steadily with increased detention time (Table 15).

TABLE 14. COMPARISONS OF CHARACTERISTICS BETWEEN REGULAR AND LIME SLUDGES OBTAINED FROM PRIMARY SEDIMENTATION OF RAW SEWAGE*

Characteristics	Lime Sludge	Regular Sludge
pH	10.5 - 11.2	7.0 - 7.5
Appearance	Light brown sludge slurry without apparent flocculent character, but slightly gelatinous.	Grayish sludge slurry with some flocculent character.
SS (%)	2.5 - 4.5	1.0 - 3.0
Volatile fraction of SS (%)	18 - 24	70 - 80
COD of VSS (mg COD/mg VSS)	1.4 - 1.5	1.4 - 1.6
Sludge settling characteristics	Very poor.	Fair to good.
Supernatant COD (mg/L)	600 - 750	300 - 400
Supernatant phosphorus (mg/L as P)		
Total-P	5 - 10	10 - 14
Ortho-P	0.1 - 0.6	8 - 12

* Huang and Tsai (1978).

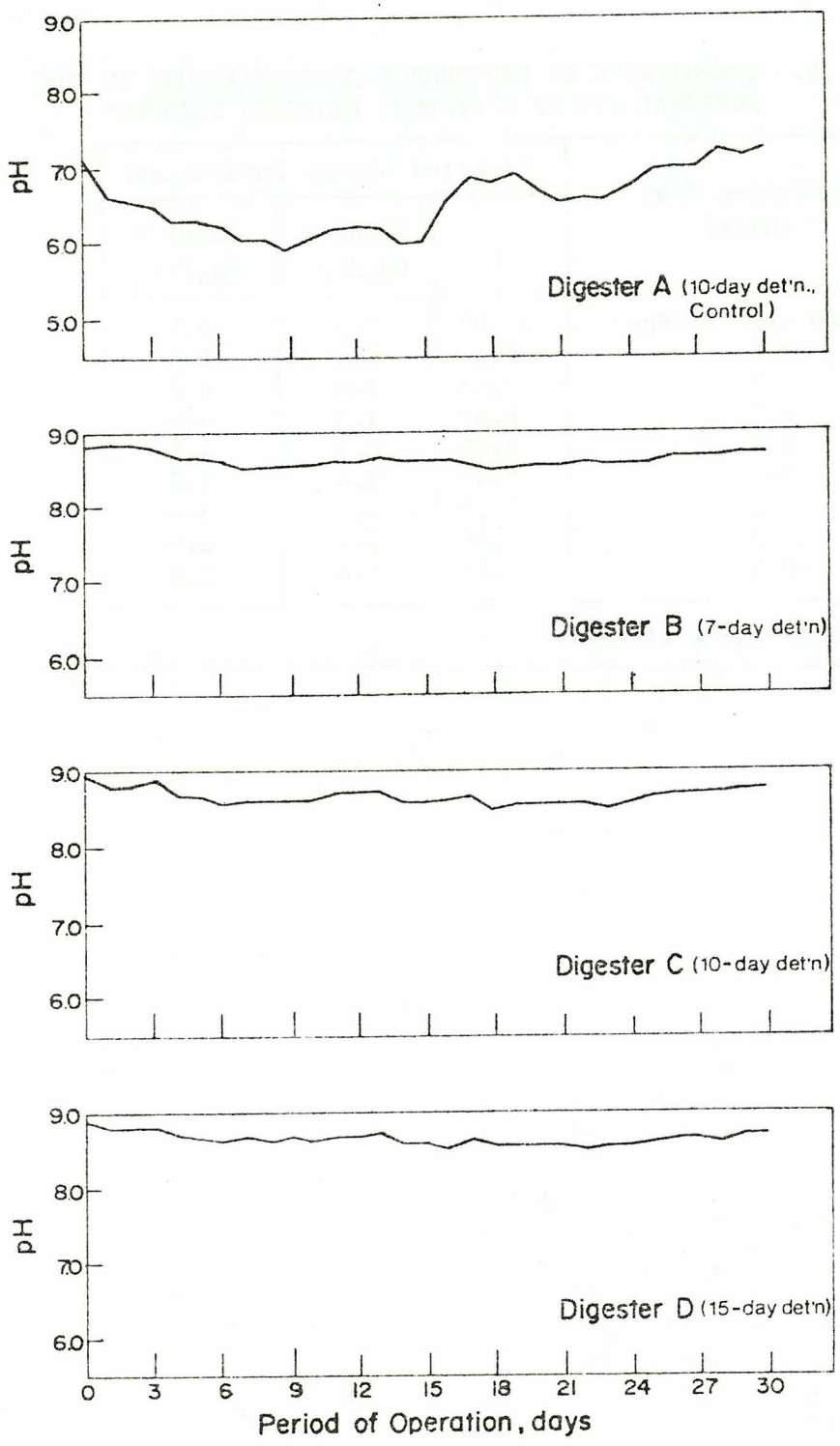
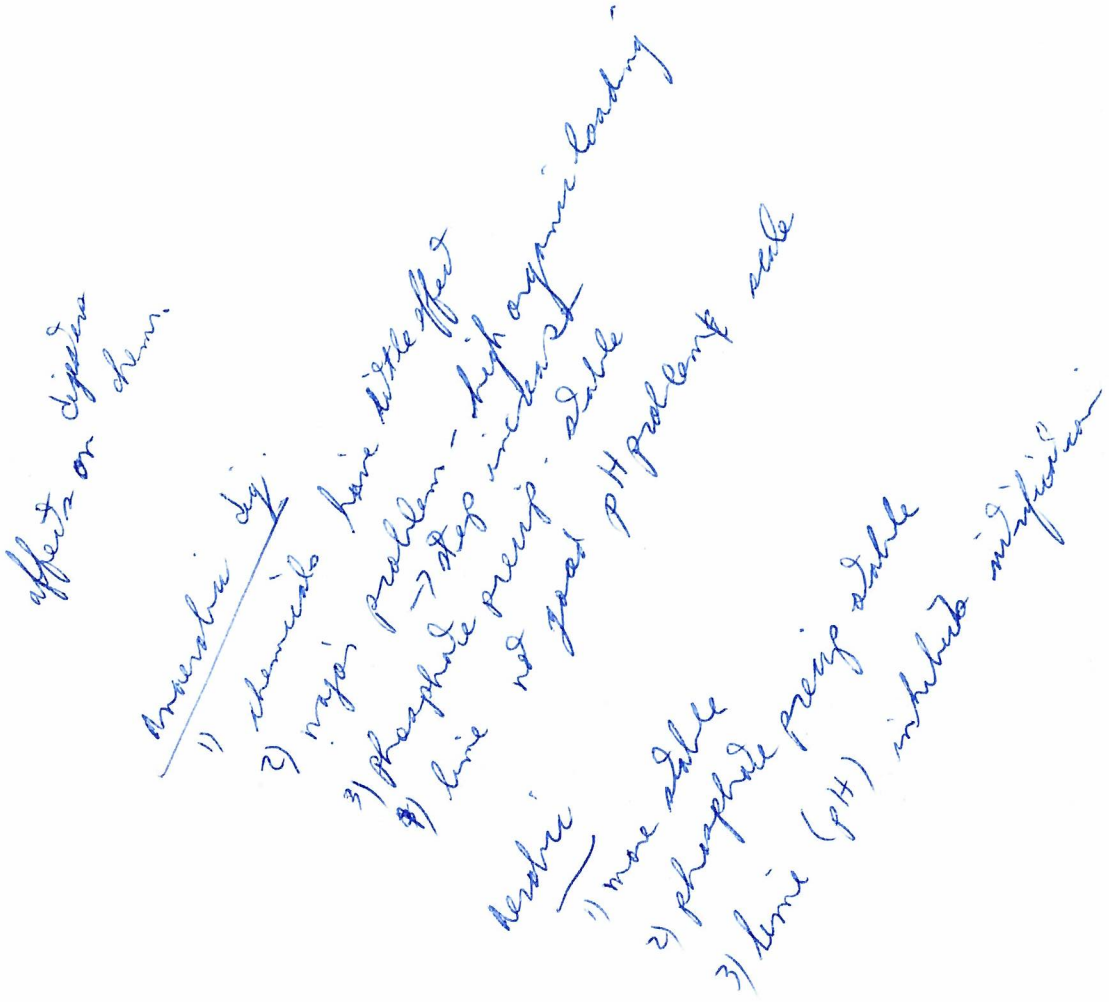


FIGURE 8: DIGESTER OPERATING DATA - pH (HAMODA AND GANCZARCZYK, 1977)

TABLE 15. COMPARISONS OF PHOSPHORUS CONCENTRATIONS IN THE SUPERNATANTS OF DIFFERENT DIGESTED SLUDGES*

Detention Time (days)	Digested Sludge Supernatant		
	pH	Ortho-P (mg/L)	Total-P (mg/L)
0 (Raw lime sludge)	11.00	0.3	6.5
1.0	9.20	1.4	4.5
1.5	8.95	1.6	4.2
2.0	8.80	1.5	4.0
2.5	8.50	1.5	4.1
3.5	8.50	1.6	3.7
5.0	8.55	1.5	3.0
10.0	8.60	1.4	2.9
20.0	8.75	1.6	2.9

* Huang and Tsai (1978).



SUMMARY

As judged from a survey of full scale plants in Ontario and from experience reported in the literature, it appears that chemical sludges resulting from phosphorus removal do not in themselves cause any upsets in anaerobic digestion systems. Any difficulties noted were generally due to a rapid overloading of the digesters following the start-up of chemical addition, a problem which may be avoided by stepwise increases in coagulant levels. Resolubilization of phosphorus from the chemical sludge during digestion was observed in only one study.

Full scale experience with lime indicated that the amount of attention required and the operational problems created leave the WPCP operators charged with responsibility for this system decidely unenthusiastic about the "lime" alternative for P-removal. This is so even though jar test results frequently indicate that lime is the least costly chemical. The indirect costs of sludge haulage and scale control added to the direct chemical costs often result in lime addition being the most costly alternative.

In view of the Ontario experience since 1971 with alum, lime and iron, it is expected that plants having to remove both phosphorus and anaerobically digest the chemically precipitated sludge, will opt in ever increasing numbers in favour of iron salts.

Although none of the surveyed aerobic digestion plants had adequate operational data, it was found that aerobic digestion is much less susceptible to upset than the anaerobic systems. A review of the literature indicated that low to moderate doses of chemical coagulants had no adverse effects on aerobic digestion, but that high dosages may upset the process. Phosphorus resolubilization from chemical sludges was insignificant. However, the high pH's associated with lime sludge had an inhibitory effect on biological nitrification.

REFERENCES

Barth, E.F. and M.B. Ettinger, "Mineral Controlled Phosphorus Removal in the Activated Sludge Process", Jour. Water Poll. Control Fed., 39:1362, 1967.

Black, S.A., "Anaerobic Digestion of Chemical Sewage Sludges, Part I - Lime Sludges", Proc. of the Technical Seminar on Aerobic & Anaerobic Sludge Stabilization Systems, Ontario Ministry of the Environment, Toronto, Ontario, 1973.

Black, S.A., "Phosphorus Removal at the Bradford WPCP", Personal communication, 1975.

Black, S.A., "Anaerobic Digestion of Lime Sewage Sludge", Canada-Ontario Agreement on Great Lakes Water Quality Research Report No. 50, Ottawa, Ontario, 1976.

Cohen, D.B., "Metro Denver's Experience with Large Scale Aerobic Digestion of Waste Activated Sludge", in Municipal Sludge Management, Proc. National Conf. on Municipal Sludge Management, Pittsburgh, Pennsylvania, June 11-13, 1974.

Cohen, D.B. and M.F.D. Hamoda, "Aerobic and Anaerobic Digestion of Chemically Precipitated Sludges", Environmental Protection Service Seminar on Alternatives for Nutrient Control, Kelowna, British Columbia, October, 1975.

Eikum, A.S., D.A. Carlson and B. Paulsrud, "Aerobic Stabilization of Primary and Mixed Primary-Chemical (Alum) Sludge", Water Research, 8:927, 1974.

Ganczarczyk, J. and M.F.D. Hamoda, "Aerobic Digestion of Organic Sludges Containing Inorganic Phosphorus Precipitates - Phase I", Canada-Ontario Agreement on Great Lakes Water Quality Research Report No. 3, Environmental Protection Service, Environment Canada, Ottawa, Ontario, June, 1973.

Gossett, J.M., P.L. McCarty, J.C. Wilson and D.S. Evans, "Anaerobic Digestion of Sludge from Chemical Treatment", Jour. Water Poll. Control Fed., 50:533, 1978.

Hamoda, M.F.D. and J. Ganczarczyk, "Aerobic Digestion of Sludges Precipitated from Wastewater by Lime Addition", Jour. Water Poll. Control Fed., 49:375, 1977.

Hsu, D.Y. and W.O. Pipes, "Aluminum Hydroxide Effects on Wastewater Treatment Processes", Jour. Water Poll. Control Fed., 45:681, 1973.

Huang, J.C. and K.C. Tsai, "New Design Criteria for Aerobic Digestion of Sewage Lime Sludge", Progress in Water Technol., 10:615, 1978.

Malhotra, S.K., T.P. Parrillo and A.G. Hartenstein, "Anaerobic Digestion of Sludges Containing Iron Phosphates", Jour. San. Eng. Div. ASCE, 97:629, 1971.

Ontario Ministry of the Environment (MOE), "1973 Operating Summary - Water Pollution Control Projects", Ontario Ministry of the Environment, Technical Services Branch, Toronto, Ontario, 1973.

Ontario Ministry of the Environment (MOE), "1974 Operating Summary - Water Pollution Control Projects", Ontario Ministry of the Environment, Technical Services Branch, Toronto, Ontario, 1974.

Shannon, E.E., J.M. Salvo and B.R. Burns, "Full Scale Phosphorus Removal Studies at CFB Petawawa", Environmental Protection Service, Technology Development Report No. EPS 4-WP-74-3, Environment Canada, Ottawa, Ontario, April, 1974.

Tofflemire, T.J. and L.J. Hetling, "Chemical-Physical Waste Treatment Phase I: The Low Lime Process", Jour. Water Poll. Control Fed., 45:210, 1973.

Van Fleet, G.L., J.R. Barr and A.J. Harris, "Treatment and Disposal of Chemical Phosphate Sludge in Ontario", Jour. Water Poll. Control Fed., 46:582, 1974.

Vesilind, P.A., Treatment and Disposal of Wastewater Sludges, Ann Arbor Science Publishers Inc., Ann Arbor, Michigan, 1974.

Zenz, D.R. and J.R. Pivnicka, "Effective Phosphorus Removal by the Addition of Alum to the Activated Sludge Process", Proc. 24th Ind. Waste Conf., Purdue University, Lafayette, Indiana, 273, 1969.

SLUDGE GENERATION, HANDLING AND DISPOSAL AT PHOSPHORUS CONTROL FACILITIES

by

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SLUDGE GENERATION, HANDLING AND DISPOSAL AT PHOSPHORUS CONTROL FACILITIES

1 INTRODUCTION

The problems of estimating sludge quantity, deciding how to handle the sludge and how to ultimately dispose of it have been with civilized man for sometime, but have been mainly ignored. The concern regarding sludge quantities, handling and disposal is generally proportional to population density, and the degree and complexity of industrialization.

For example, consider for a moment that even when going back to early biblical records, nowhere does it mention that in the construction of the ark Noah considered, or even anticipated the monumental sludge handling problem he would have to face once he had all his animals on board. His solution when faced with the problem once adrift, is left to your imagination.

It would appear that still too frequently design engineers today suffer from the "Noah Syndrome".

Early man considered sludge as a resource, something to be recycled. Many less industrialized nations still pursue this philosophy. Even, highly industrialized nations, not blessed with an abundance of resources have prescribed to a similar philosophy. In North America, it took an energy crisis to redirect our thinking to the point where sludge is looked at from a utilization rather than disposal perspective.

It has been said that "history repeats itself". This then would also appear to be true when dealing with sludge.

This paper will focus on providing information on the effect of adding metal salts to existing wastewater treatment plants for phosphorus removal to 1 mg/L total as it impacts on sludge quantity, handling and disposal/utilization.

The information is based on data from 185 waste treatment plants surveyed in the Province of Ontario. The completeness of data varies considerably.

2 SLUDGE QUANTITIES

Waste treatment process design engineers are continually plagued by lack of information when it comes to designing sludge handling and disposal/utilization facilities. The length of the short cut to data acquisition is pretty well proportional to the degree of confidence to be placed in the process capacity design.

Good data is hard to get and costs money. The original error made in sludge quantity estimation can and is increased when attempting to estimate resulting sludge quantities due to chemical addition for phosphorus removal requirements. This error becomes greater as the degree of phosphorus removal increases from a target of 1 $mg \cdot L^{-1}$ to 0.1 $mg \cdot L^{-1}$, and can be attributed to increasingly greater deviations from stoichiometric relationships between influent P and effluent target P.

Sludge production is influenced by a number of variables. For chemical sludge production, the chemical used, wastewater characteristics and point of chemical addition all play an important role. For biological sludge production, the type of process used in the conversion of substrate greatly affects the amount of biomass produced. The amount of sludge produced also varies with the nature of substrate oxidized. Higher sludge volumes result in winter than in summer because the auto-oxidation rate depends on temperature. The total volume of sludge produced from biological and physical/chemical systems, or any combination thereof is also influenced by clarifier performance, sludge recycle and the degree of operator attention to the system.

The literature abounds with sludge production data. Figure 1 is just one example for municipal sludges and illustrates the degree of

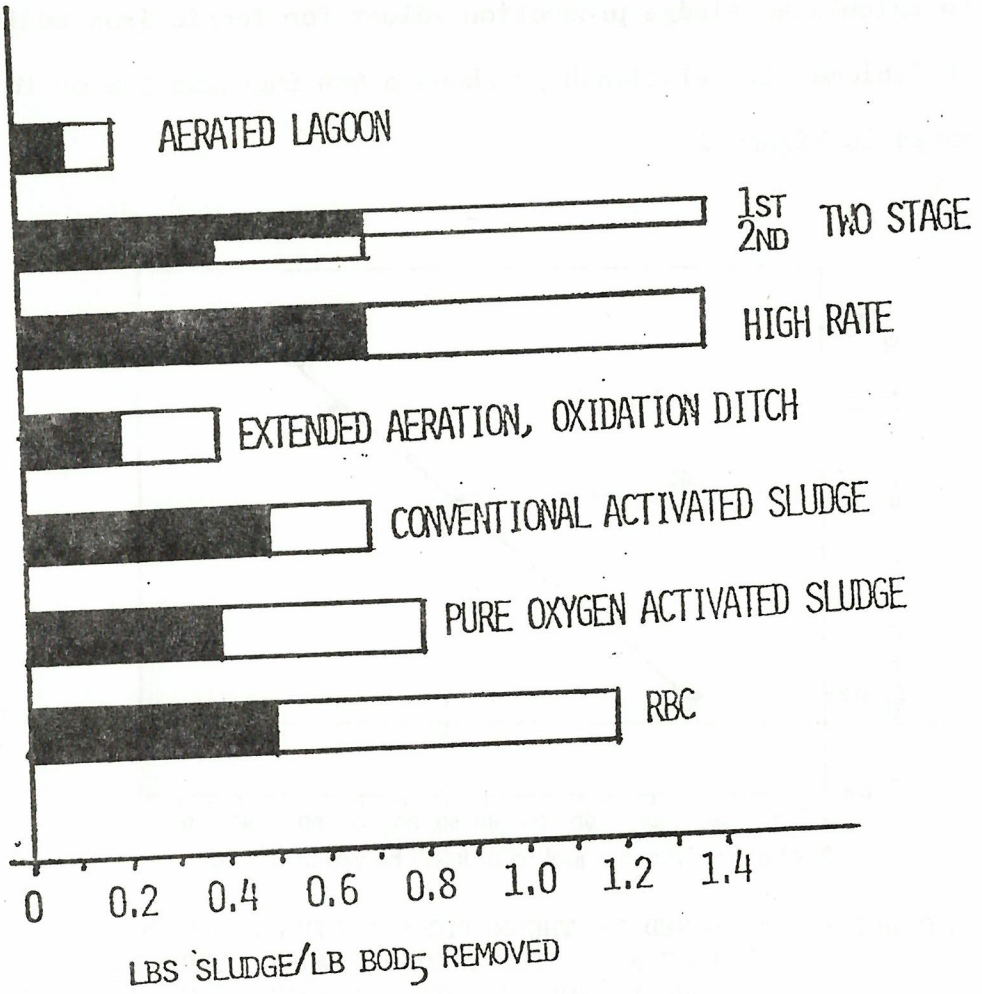


FIGURE 1. BIOLOGICAL SLUDGE PRODUCTION [1]

variability which can be in excess of 100%.

A most useful method of sludge quantity estimation consists of performing a mass balance around various treatment process components and coupling this with process efficiency assumptions (2).

Calculations to determine chemical sludge quantities based on stoichiometric relationships have been illustrated by Campbell (3).

Pilot Scale Activated Sludge P and N Removal Studies

A long-term pilot scale study for the removal of phosphorus and nitrogen (4) was conducted at the Wastewater Technology Centre (WTC). Sludge production was monitored over a 16-day consecutive period. Comparing the

observed to calculated sludge production values for ferric iron addition based on stoichiometric relationships shows a 65% increase 50% of the time, as illustrated in Figure 2.

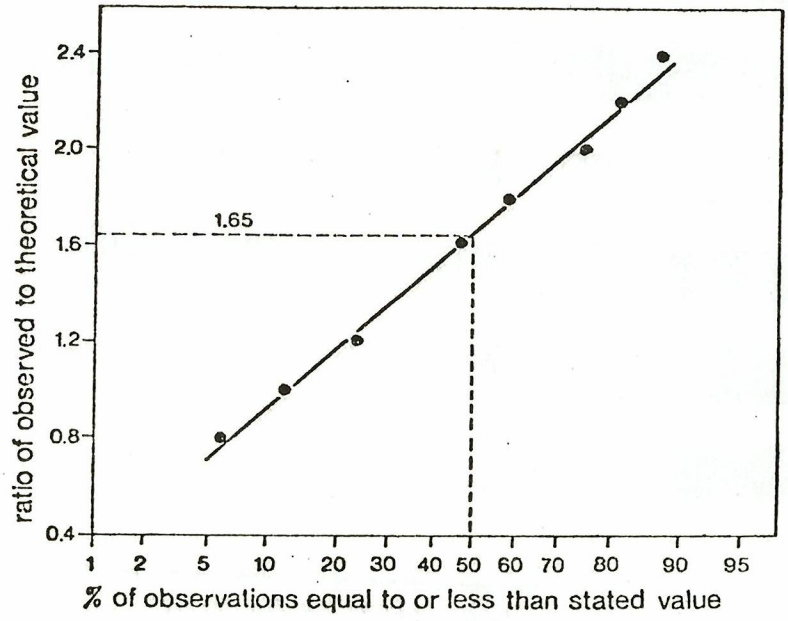


FIGURE 2. OBSERVED TO THEORETICAL SOLIDS PRODUCTION RATIO-FERRIC IRON ADDITION AT A NITROGEN AND PHOSPHORUS REMOVAL ACTIVATED SLUDGE PLANT[4]

Full Scale P-Removal Studies - Primary Plant

Full scale phosphorus removal studies were conducted at the primary wastewater treatment plant at C.F.B. Borden (5). The study lasted ten months and covered three phases of chemical addition for phosphorus removal using lime, alum and ferric chloride. While the major objective of this study was to determine the optimum phosphorus removal precipitant and its dosage to achieve an effluent P objective of $\leq$ 1 mg/L, information on sludge production under various operational conditions was also collected. These data as summarized in Table 1 were compared to calculated sludge production values and are shown as a frequency distribution in Figure 3. In this case, the amount of sludge produced was overestimated by 28%, 50% of the time.

TABLE 1. CAMP BORDEN SLUDGE PRODUCTION - PRIMARY PLANT[5]

P-Removal Precipitant		Sludge Mass Produced				Ratio of Measured to Calculated Value
		Calculated		Measured		
Chemical	Dosage	kg/m³	lbs/10⁶gal	kg/m³	lbs/10⁶gal	
-	mg/l					
Baseline	-	116	255	191	420	1.65
Lime	151[1]	1186	2609	1148	2526	0.97
	197	1389	3056	1400	3080	1.01
	275	3184	6426	1697	3425	0.53
	210	1876	3857	1399	2876	0.76
Alum	4.4[2]	201	520	332	859	1.65
	7.5	385	1020	283	750	0.74
	14.8	645	1730	253	679	0.39
	18.5	604	1601	294	779	0.49
Ferric Chloride	9.6[3]	290	760	304	796	1.05
	14.6	321	872	307	834	0.96
	19.0	502	1331	312	827	0.62
	26.6	515	1382	343	920	0.67

[1] as $Ca(OH)_2$

[2] as Al^{3+}

[3] as Fe^{3+}

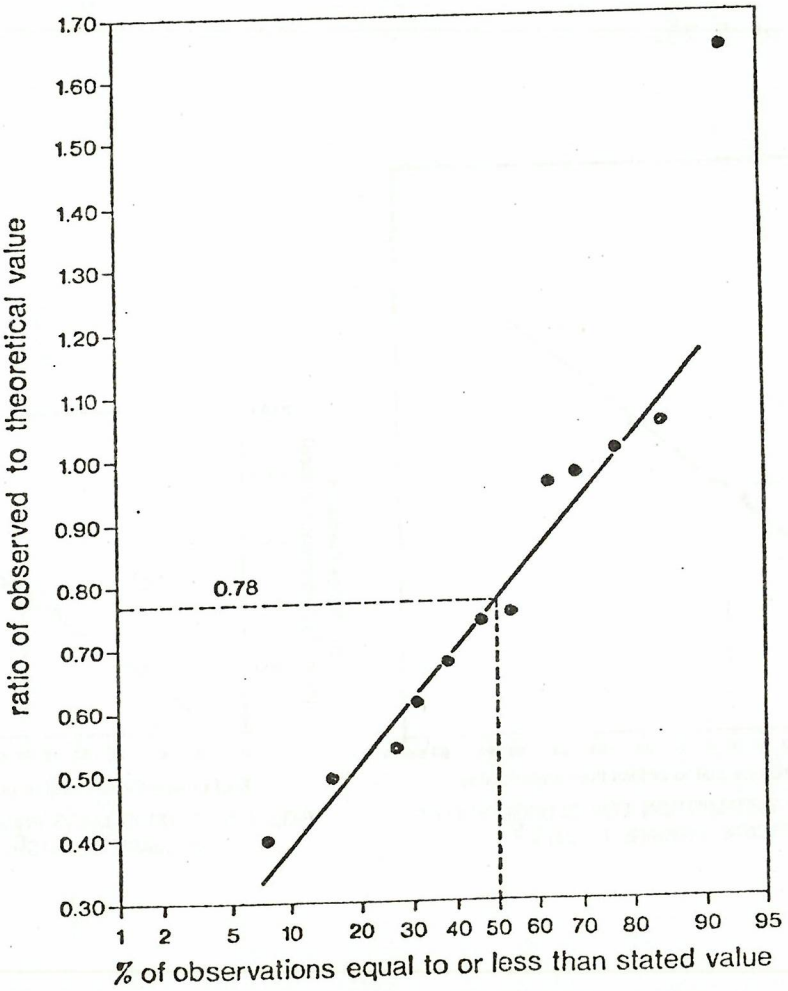


FIGURE 3. OBSERVED TO THEORETICAL SOLIDS PRODUCTION RATIO-PRECIPITANT ADDITION TO A FULL SCALE PRIMARY PLANT [5]

Ontario Treatment Plant Survey Data

Sludge production data were obtained in a 1975 survey of Ontario wastewater treatment plants where records prior to phosphorus removal were compared with plant records following installation of phosphorus precipitation systems (6).

Figure 4 summarizes the data from 15 conventional, primary plants (without precipitant addition) surveyed covering a range of hydraulic loadings from 0.26 to 11 MGD*. The data show that 50% of time 1 995 gallons of sludge are produced for each million gallons of wastewater treated. This translates to 1 140 lbs dry solids for each million gallons treated (Figure 5). The total solids concentrations of the raw primary sludges varied from 3.5 to 8% with a mean of 5.7%.

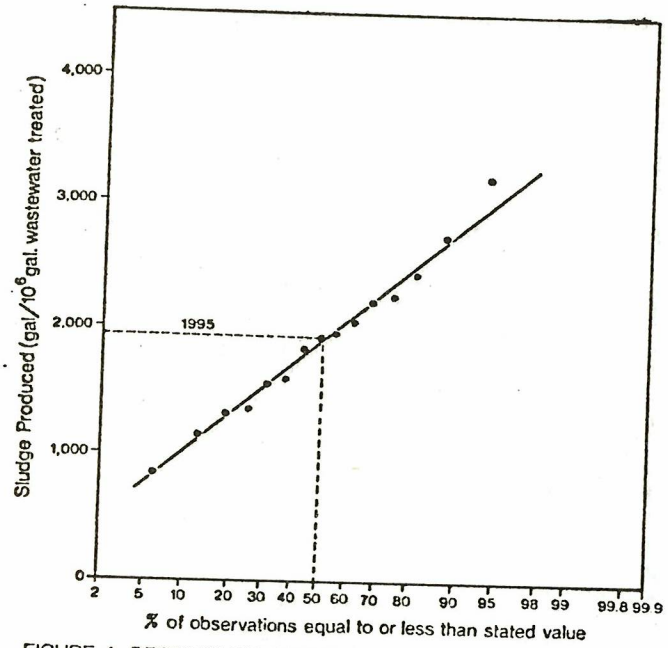


FIGURE 4. PROBABILITY DISTRIBUTION FOR SLUDGE VOLUME PRODUCED AT CONVENTIONAL PRIMARY PLANTS [6]

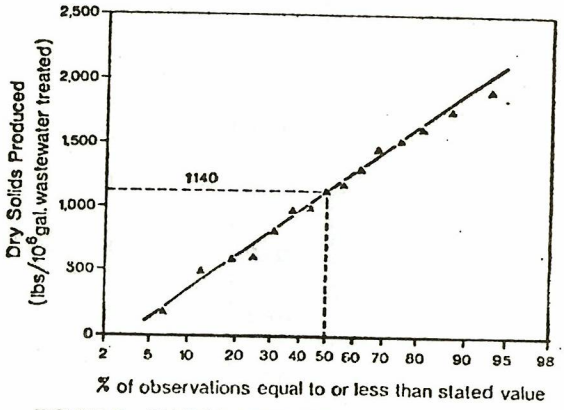


FIGURE 5. SLUDGE MASS PRODUCED AT CONVENTIONAL PRIMARY PLANTS [6]

*Imperial gallons used throughout this paper

The impact of chemical addition for phosphorus removal at primary plants is illustrated in Figure 6 for seven upgraded plants. In these plants, the average sludge solids concentration decreased from 6.0 to 5.3% after chemical addition. The sludge mass increased by 40%.

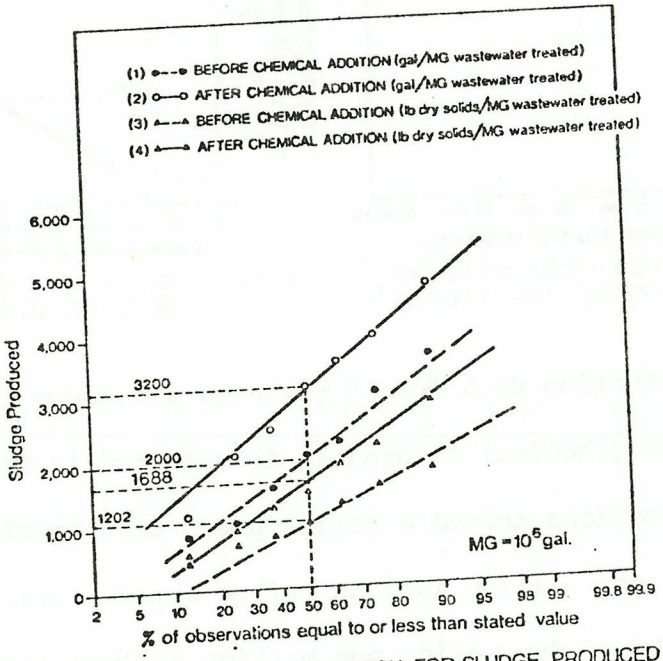


FIGURE 6. PROBABILITY DISTRIBUTION FOR SLUDGE PRODUCED AT PRIMARY PLANTS WITH ADDITION OF METAL SALTS 5

Sludge production data from 42 secondary plants using the conventional activated sludge process were analyzed. The plants have flow capacities ranging from 0.3 to 170 MGD. The raw sludge produced consists of both primary and waste activated sludge. In the case of conventional activated sludge plants, Figure 7 shows that 50% of the time at least 3 905 gallons of sludge are produced per million gallons treated. Solids concentrations varied from 2 to 7%, with a weighted average of 4.6%. Similarly, the dry weight of solids produced at conventional activated sludge plants was equal to or less than 1 786 lbs dry solids per million gallons of wastewater treated, 50% of the time (Figure 8).

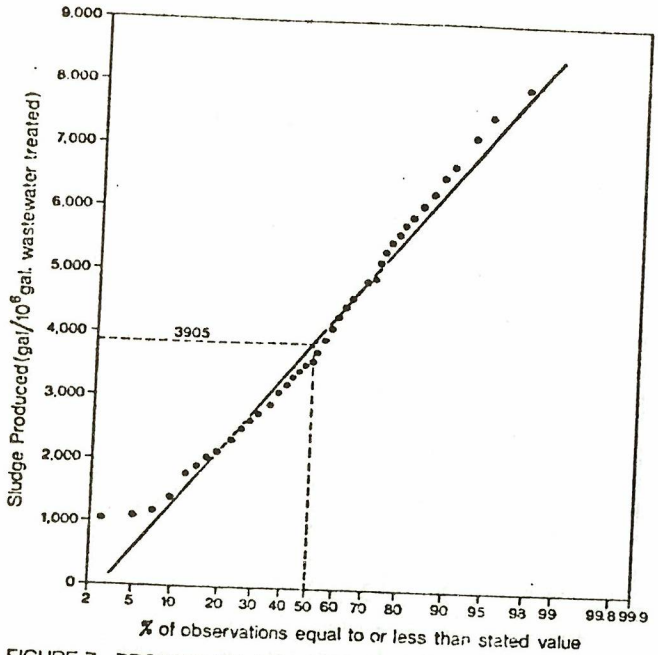


FIGURE 7. PROBABILITY DISTRIBUTION FOR SLUDGE VOLUME PRODUCED AT CONVENTIONAL SECONDARY (C.A.S.) PLANTS [6]

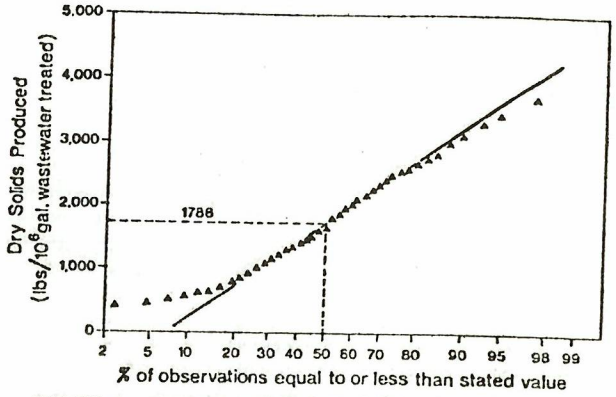


FIGURE 8. SLUDGE MASS PRODUCED AT CONVENTIONAL SECONDARY (C.A.S.) PLANTS [6]

Sludge production data for 15 upgraded secondary plants (primary and waste activated, chemical sludge) is illustrated in Figure 9. Fifty percent of the observations showed a solids production equal to or less than 1 725 lbs dry solids per million gallons before chemical addition. This increased to 2 175 lbs of dry solids per million gallons after chemical addition and represents a 26% increase in sludge mass. Following precipitant addition, the average total solids concentration decreased from 4.5 to 4.2%.

While metal salts are generally added to the aeration tanks, data analyzed from four installations where metal salts were added to the primary settling tank showed a decrease in solids produced. In this instance, the lower organic loading to the aeration tank due to additional organics removed in the primary, resulted in reduced biosynthesis.

Summary of Sludge Production

Sutton (4) underestimated sludge production by 65% when using stoichiometric relationships for a biological system with chemical addition. Stepko (5) however, overestimated sludge production resulting from chemical addition to a primary plant by 28%. These studies exemplify the problems associated with estimating sludge production from chemical stoichiometry.

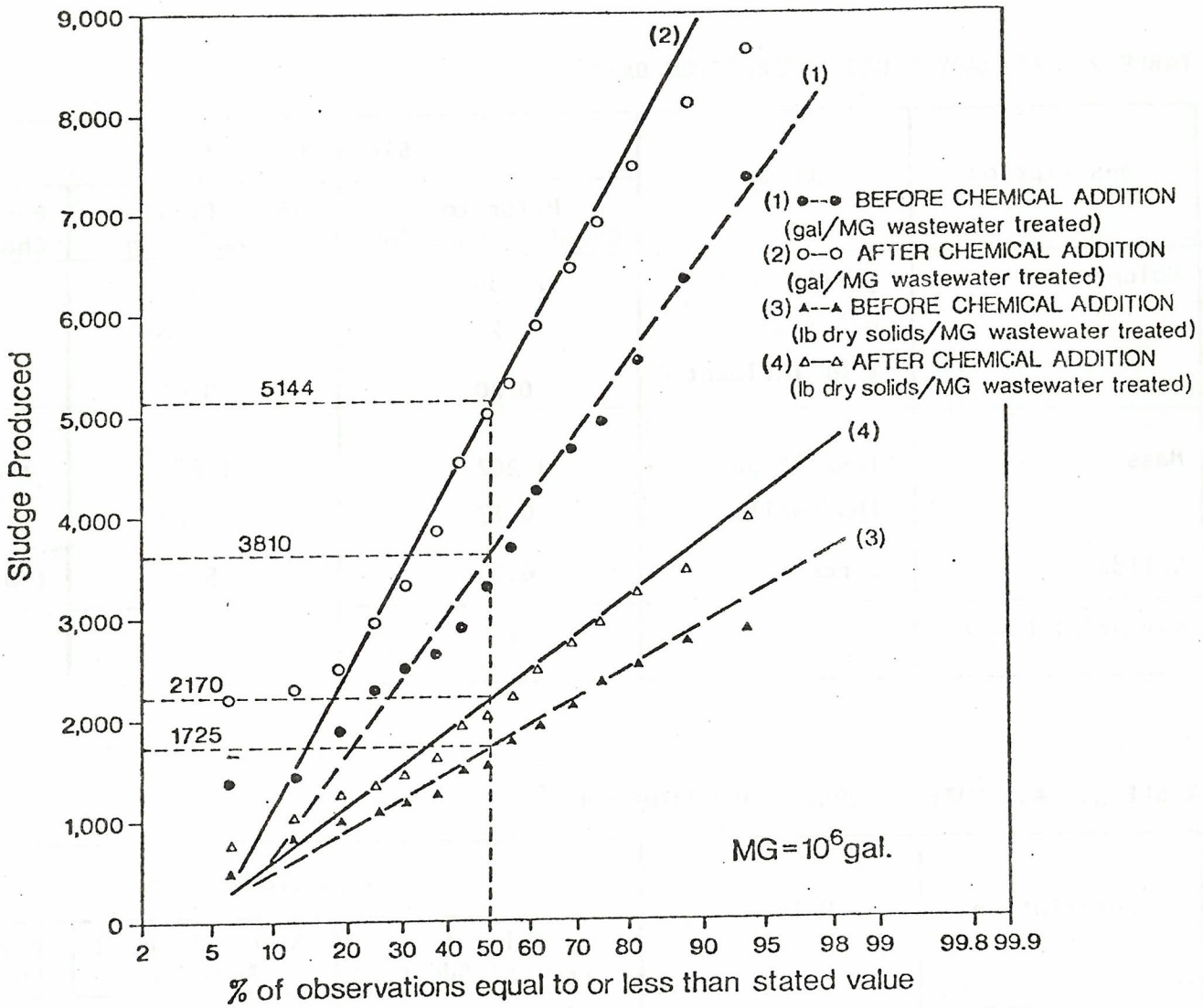


FIGURE 9. PROBABILITY DISTRIBUTION FOR SLUDGE PRODUCED AT SECONDARY (C.A.S.) PLANTS WITH ADDITION OF METAL SALTS TO AERATION TANK [6]

From our experience, the best data base for sludge production exists in the Ontario survey of full scale treatment plants (6). The data presented in earlier figures are summarized in Tables 2 and 3 for primary and activated sludge plants, respectively.

TABLE 2. PRIMARY SLUDGE PRODUCTION DATA[6]

Description	Units	Sludge Production		
		Prior to Chemical Addition	After Chemical Addition	Percent Change
Volume	gal/10^6 gal	2 000	3 200	+ 60
	gal /capita	0.29	0.46	-
	% of influent Q	0.20	0.32	-
Mass	lbs/10^6 gal	1 202	1 688	+ 40
	lbs/capita	0.17	0.24	-
Solids	percent	6.0	5.3	|-0.7|
Number of Plants	-	7	7	-

TABLE 3. ACTIVATED SLUDGE PRODUCTION DATA [6]

Description	Units	Sludge Production		
		Prior to Chemical Addition	After Chemical Addition	Percent Change
Volume	gal/10^6 gal	3 810	5 144	+ 35
	gal/capita	0.55	0.75	-
	% of influent Q	0.38	0.51	-
Mass	lbs/10^6 gal	1 725	2 175	+ 26
	lbs/capita	0.25	0.32	-
Solids	percent	4.5	4.2	|-0.3|
Number of Plants	-	15	15	-

Based on the results of the Ontario survey (6), some generalizations concerning sludge production design data are shown in Table 4.

TABLE 4. SLUDGE PRODUCTION - SUGGESTED DESIGN DATA*

System	lbs pcd	Sludge Quantity	
		Volume % of Influent	lbs d.s./10^6 gal
Conventional Primary	0.17	0.20	1 200
Upgraded Primary	0.24	0.32	1 700
Conventional A.S**	0.25	0.38	1 725
Upgraded A.S.**	0.32	0.51	2 175

* based on Q = 145 gpcd
d.s. = dry solids
pcd = per capita/day

** primary + waste activated

The rule-of-thumb that sludge volume approaches 0.5% of the influent hydraulic load to a conventional plant is a good approximation. By using this estimate, the apparent margin of safety would allow upgrading of a conventional plant to include chemical phosphorus removal to 1.0 mg/L total phosphorus using metal salts without major expansion of sludge handling facilities.

Because few Ontario plants practice P removal using lime, no substantive data base for sludge quantity estimation exist. However, based on past experience at a number of pilot and full scale facilities practicing P removal using lime, reasonable estimates of sludge production can be made.

The mass of sludge produced will depend largely on the wastewater alkalinity and the lime dosage required to attain a specific pH at which the target P effluent level is achieved.

Figure 10 illustrates that, having determined the pH at which the P effluent target will be achieved, the correlation indicates the lime/alkalinity ratio required. Knowledge of the wastewater alkalinity enables calculation of the required lime dosage (7). Another correlation (9) for raw wastewaters from 20 Ontario municipalities showing alkalinity/lime dosage requirements to attain pH 10 and 11 is shown in Figure 11.

Sludge Quantities After Anaerobic Digestion

The sludge production data summarized earlier, facilitates the design of sludge handling and volume reduction facilities. When designing facilities for ultimate disposal, the sludge volume after anaerobic digestion must be known. Such data are difficult to obtain. In many instances, this can be attributed to incomplete records concerning volume of sludge disposed of, as well as problems associated with solids concentration determinations.

The Ontario survey (6), while incomplete, provides the best, currently available data base on this subject. The data relating sludge volumes disposed from standard primary plants to population served were subjected to regression analysis (Figure 12).

The equation expressing this relationship for 17 plants is shown as:

$$\text{Sludge Disposed} = 0.0169\ (\text{Population} \times 10^{-3})^{1.131} \quad (\times 10^6\ \text{gal/yr}) \qquad (1)$$

Figure 12 illustrates the fact that digester problems will result in substantial increases in sludge volumes requiring disposal.

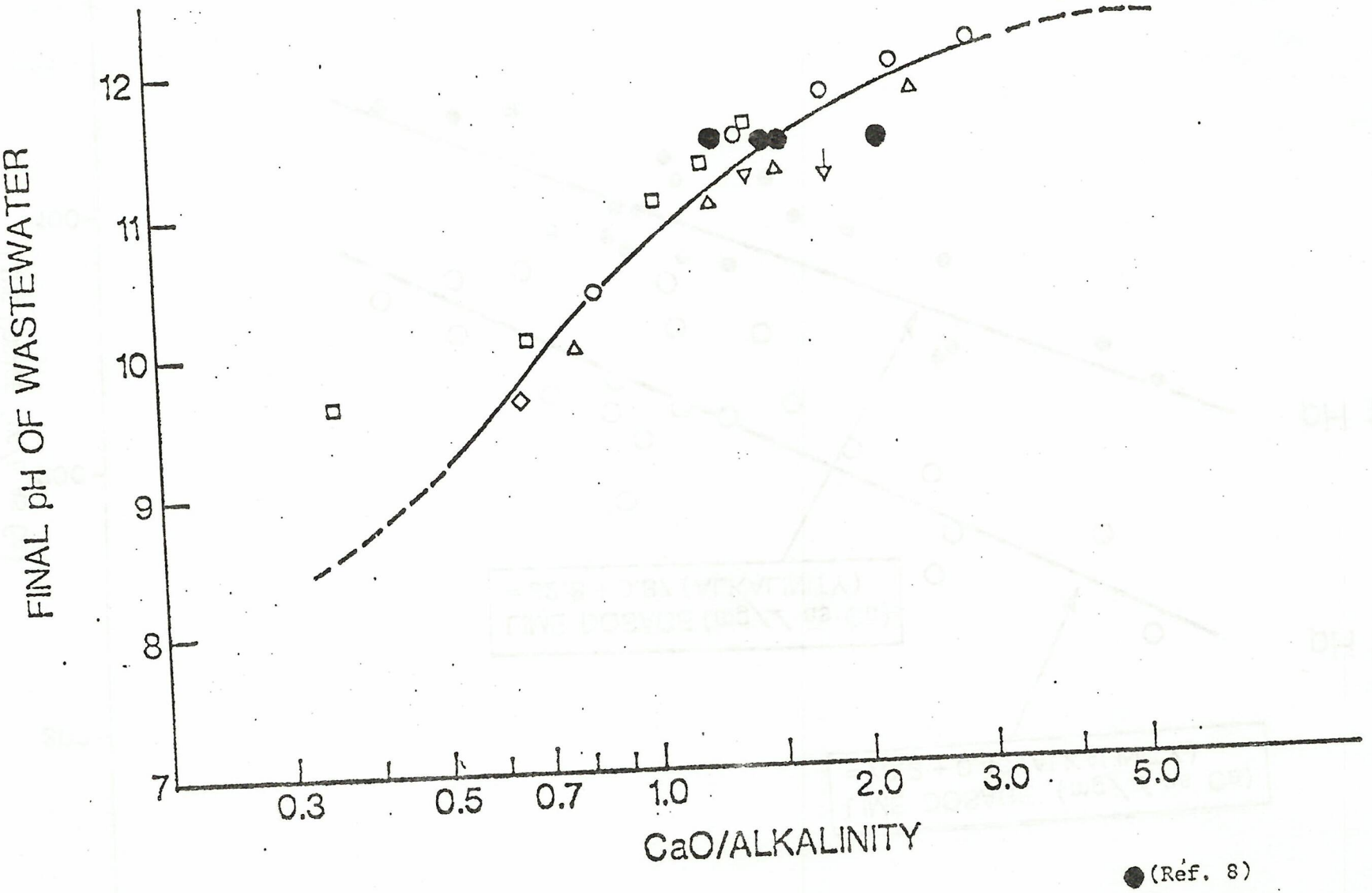


FIGURE 10. RATIO OF LIME DOSAGE (MG/L) TO INITIAL WASTEWATER ALKALINITY[7]

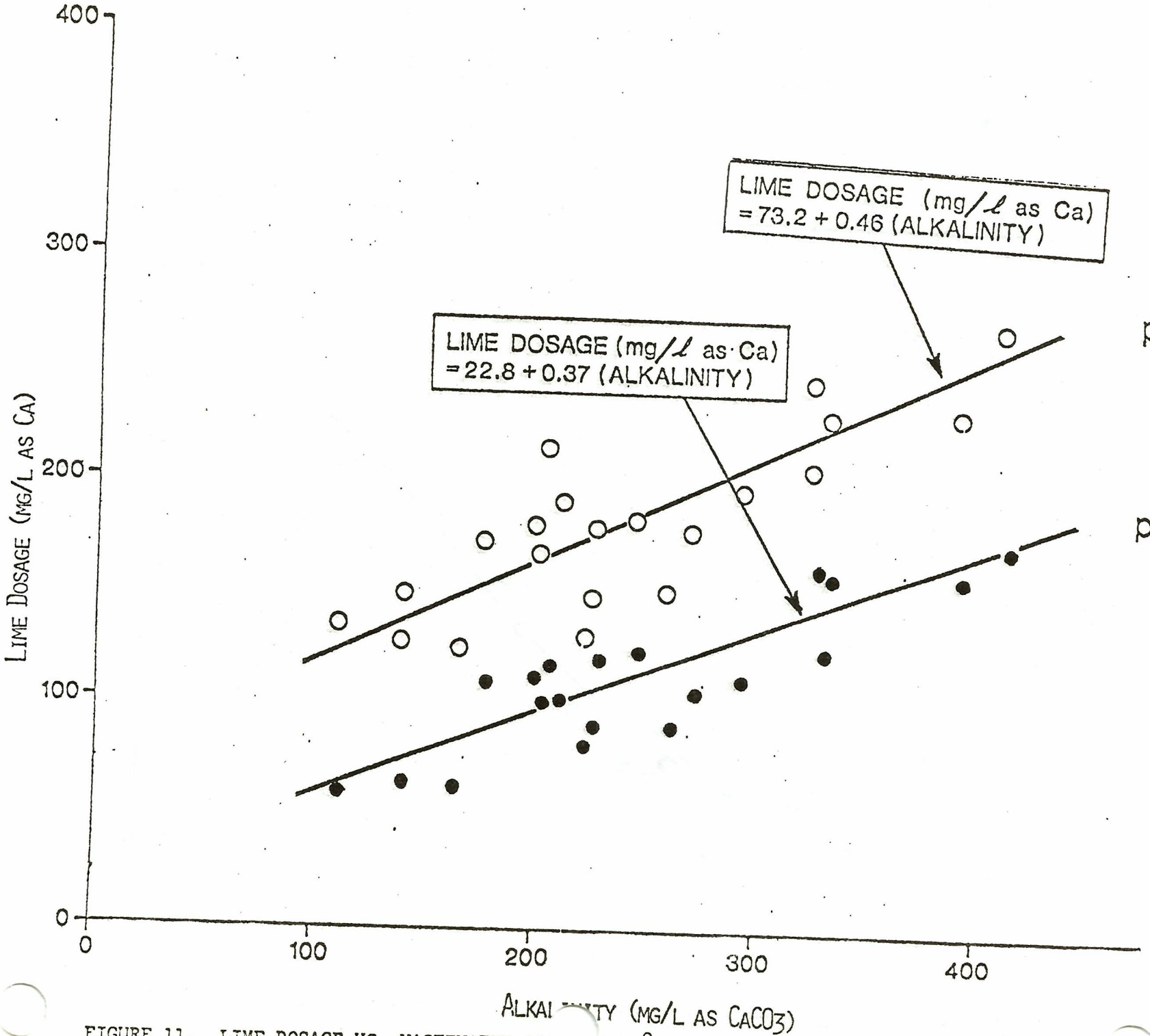


FIGURE 11. LIME DOSAGE VS. WASTEWATER ALKALINITY [9]

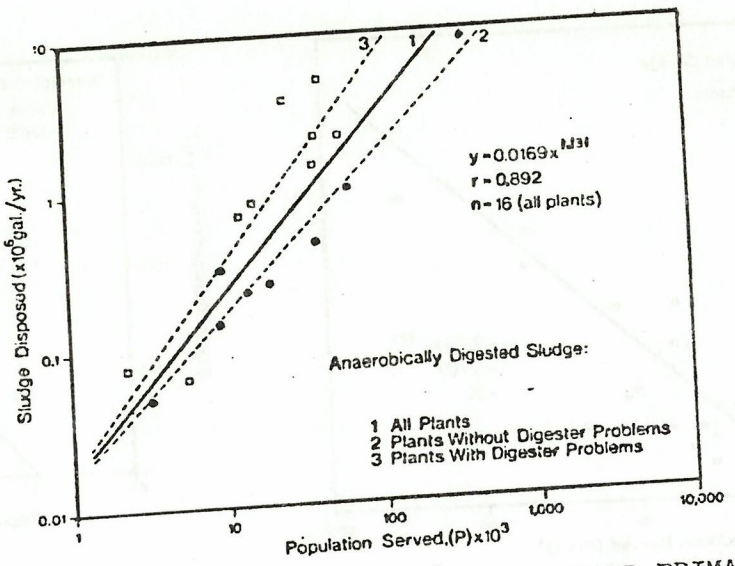


FIGURE 12. SLUDGE DISPOSAL AT STANDARD PRIMARY PLANTS-SLUDGE VOLUME VS. POPULATION SERVED [6]

Similarly, Figures 13 to 16 illustrate from the available Ontario data (6) various relationships between volume or mass of sludge to be disposed of from activated sludge plants after anaerobic digestion as a function of population served. The equations are summarized in Table 5.

TABLE 5. SUMMARY OF ANAEROBICALLY DIGESTED SLUDGE DISPOSAL—RELATIONSHIPS WITH POPULATION SERVED[6]

Standard C.A.S. Plants		
n = 22	Sludge disposed ($\times 10^6$ gal/yr) = 0.131 (Population $\times 10^{-3}$)$^{0.873}$	(2)
n = 23	Sludge disposed (TS, ton/yr) = 11.53 (Population $\times 10^{-3}$)$^{1.097}$	(3)
Upgraded C.A.S. Plants		
n = 22	Sludge disposed ($\times 10^6$ gal/yr) = 0.105 (Population $\times 10^{-3}$)$^{1.089}$	(4)
n = 23	Sludge disposed (TS, ton/yr) = 13.71 (Population $\times 10^{-3}$)$^{1.146}$	(5)

Figure 17 summarizes all the pertinent water pollution control plant data from the Ontario survey (6).

Sludge Quantity Predictions for Lower Than 1 $mg \cdot L^{-1}$ Effluent P Targets

The aforenoted information does not address the question of "how much more sludge would be generated when imposing point source controls for effluent total phosphorus concentrations of 0.5 $mg \cdot L^{-1}$ or even 0.1 $mg \cdot L^{-1}$?".

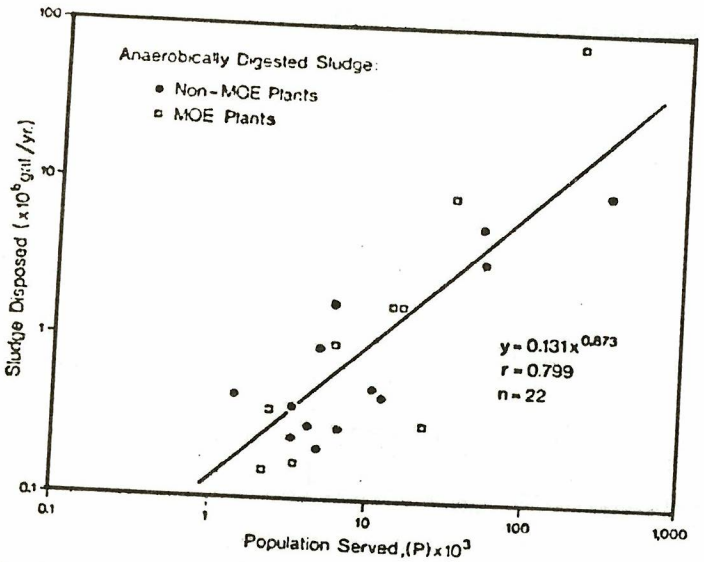


FIGURE 13. SLUDGE DISPOSAL AT STANDARD C.A.S. PLANTS-SLUDGE VOLUME VS. POPULATION SERVED 6

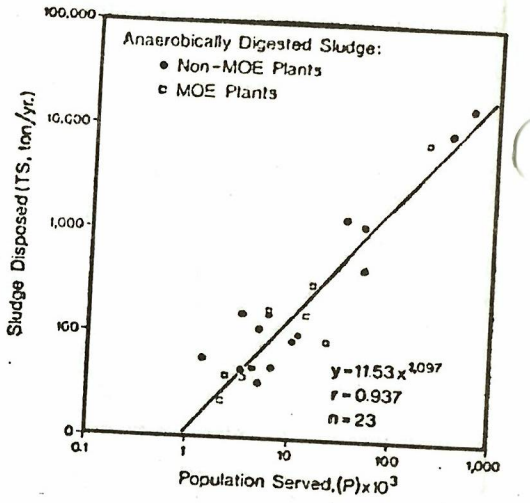


FIGURE 14. SLUDGE DISPOSAL AT STANDARI C.A.S. PLANTS-DRY WEIGHT OF SLUDGE VS. POPULATION SERVE

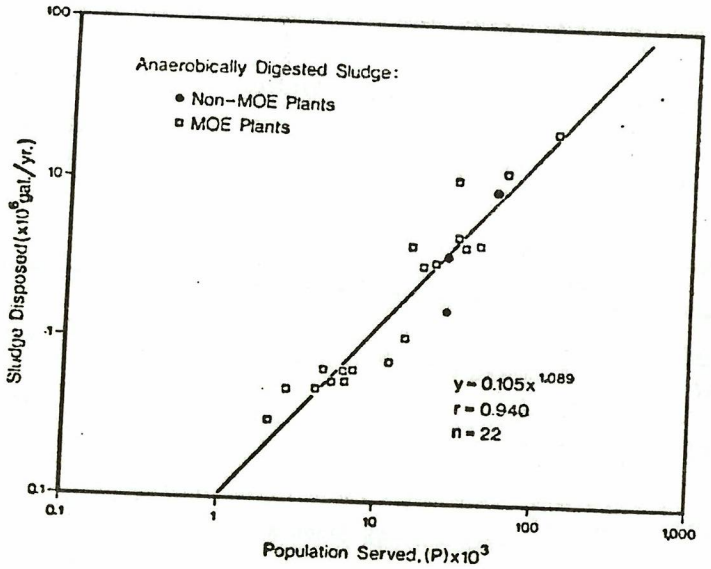


FIGURE 15. SLUDGE DISPOSAL AT UPGRADED C.A.S. PLANTS-SLUDGE VOLUME VS. POPULATION SERVED 6

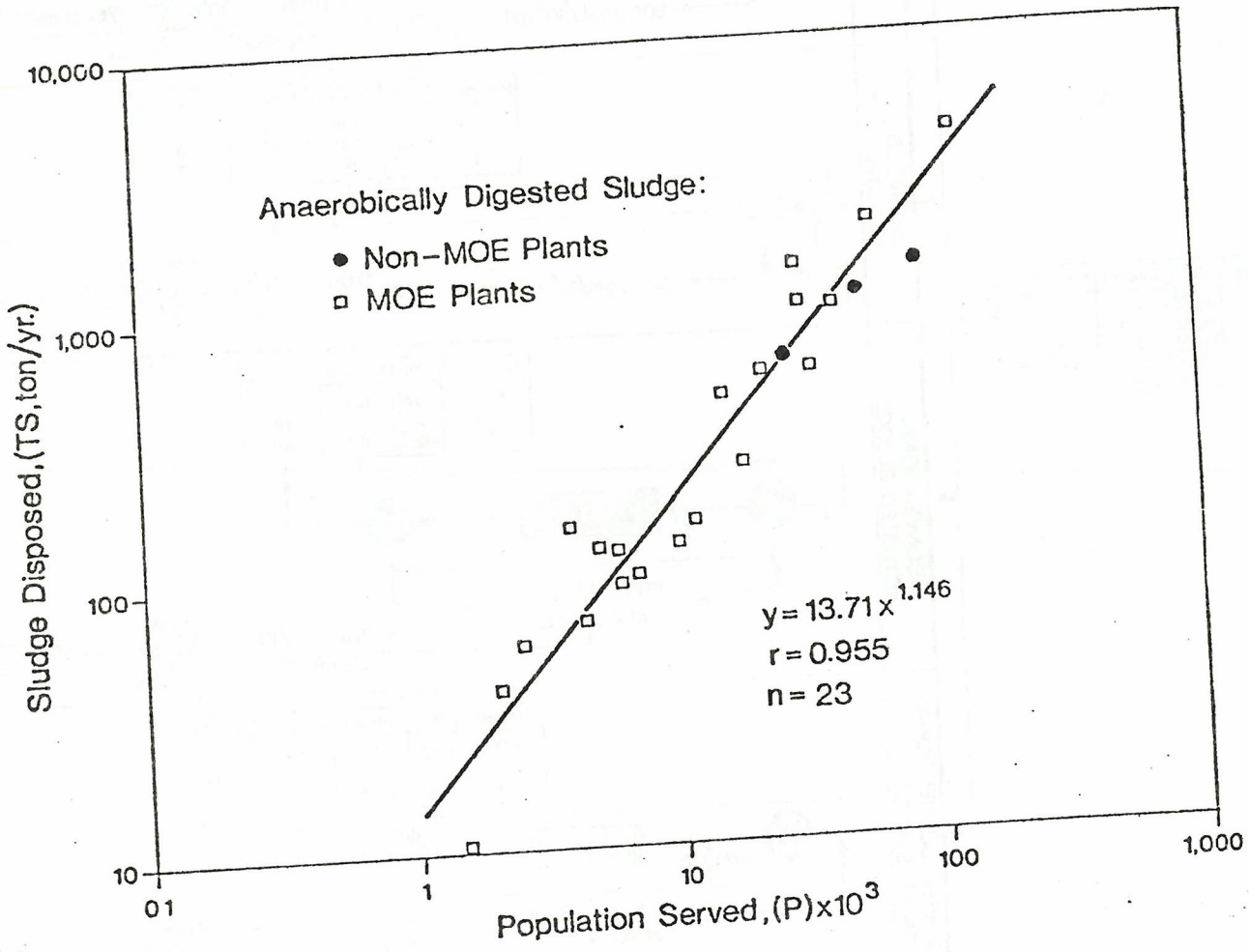


FIGURE 16. SLUDGE DISPOSAL AT UPGRADED C.A.S. PLANTS-DRY WEIGHT OF SLUDGE VS. POPULATION SERVED[6]

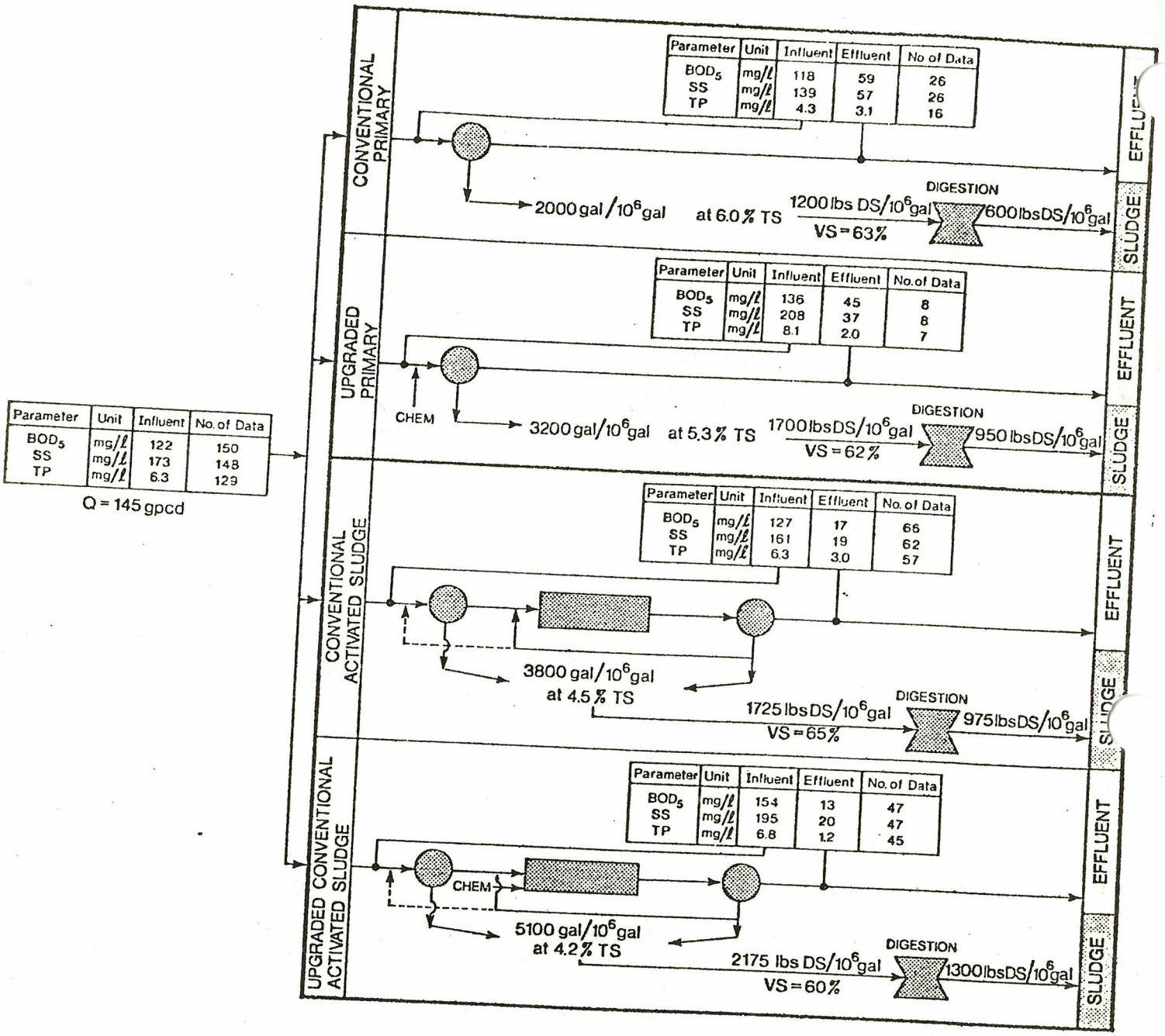


FIGURE 17. SUMMARY OF ONTARIO WATER POLLUTION CONTROL PLANT SLUDGE PRODUCTION AND DISPOSAL DATA[6]

A recent document (10) made a first attempt at answering this question by reporting on a computer simulation of required process modifications to meet various point source P control scenarios and the resulting sludge quantities. Figure 18 is a typical illustration of the dramatic increases in sludge mass over baseline conditions of no phosphorus removal. The example shows a simulation for 17 Canadian plants in the Lake Ontario drainage basin and represents a total flow of 582 MGD for a sewered population of 3.8 million persons. The simulation predicts a sludge mass increase from 34% over baseline conditions (no chemical addition) for an effluent total phosphorus target concentration of 1.0 $mg \cdot L^{-1}$.

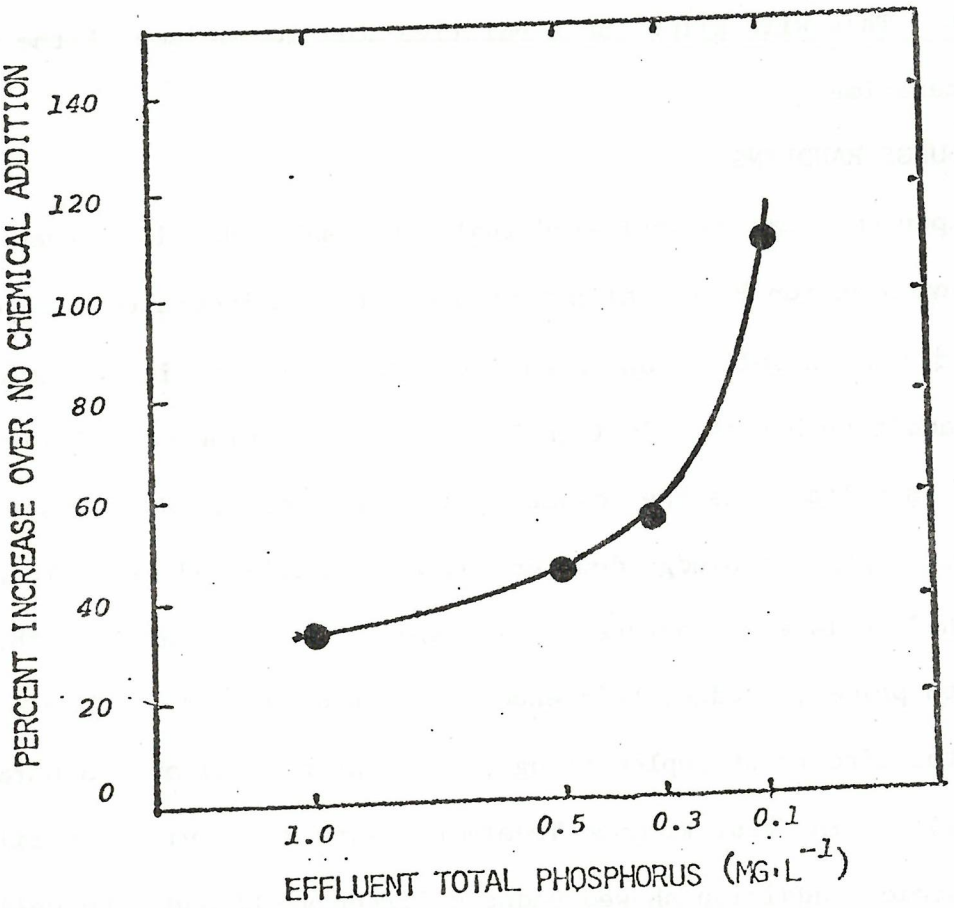


FIGURE 18. INCREASE IN SLUDGE MASS DUE TO CHEMICAL ADDITION TO MEET VARIOUS EFFLUENT TOTAL PHOSPHORUS TARGETS

This prediction compares favourably with the 1975 Ontario sludge survey data (6) indicating a 26% sludge mass increase. Treatment process modifications required to attain an effluent objective of 0.1 $mg \cdot L^{-1}$ total phosphorus predicts a 108% increase in sludge mass over baseline levels (no chemical addition).

It is the intent of the Phosphorus Management Strategy Task Force to update and refine the prediction model and input available data for U.S. and Canadian plants in the Great Lakes Basin. This will then enable the Task Force to identify the impact of various P control point source control scenarios at municipal plants on sludge quantities that might be generated.

The updated version of the model will also generate capital as well as O&M costs. This will allow for a relative cost comparison of the various P control scenarios.

SLUDGE HANDLING

Experience has demonstrated that metal salt addition to wastewater treatment processes for P removal not only results in increased sludge volumes and mass, but reduced solids concentration. The increased inorganic content due to chemical addition has the additional effect of lowering calorific values if incineration is selected as the volume reduction process. More ash is produced.

With regard to sludge dewatering, waste activated alum sludges, because of their gelatenous nature, are generally not dewatered by themselves but mixed with primary sludge, thickened and then dewatered. Table 6 illustrates the effects of implementing phosphorus removal at two Ontario treatment plants. The West Windsor Treatment Plant is a primary facility which upon chemical addition showed reduced filter yield and cake solids concentration. Filtrate solids and conditioning costs increased. These effects were more pronounced with alum than ferric chloride. The North

TABLE 6. FULL SCALE VACUUM FILTRATION OF SLUDGES FROM PHOSPHORUS REMOVAL FACILITIES

Description	West Windsor[1]			North Toronto[1]	
Phosphorus Removal Chemical	None	Fe^{3+}	Al^{3+}	None	Fe^{3+}
Type of Sludge	primary			digested elutriated	
Solids Concentration (%)	11.9	8.6	7.9	8.1	7.6
Conditioning Chemicals (% lime) (% ferric chloride)	9.9 1.3	15.9 0.1	24.0 1.2	9.5 0	11.8 6.6
Conditioning Cost[2] ($/ton of dry solids)	4.01	6.69	7.39	Not Recorded	Not Recorded
Filter Yield ($lbs/ft^2.h$)	12.4	9.6	6.7	3.8	3.9
Filter Cake Solids (%)	31	21	17	23	19
Filtrate SS (mg/L)	2 830	3 660	13 900	5 550	7 690

1 Campbell et al (11)
2 Cost Figures as of December, 1975

Toronto Sewage Treatment Plant experience using ferric chloride also showed a decrease in filter cake solids and increases in sludge conditioning requirements. No decrease in filter yield was noted.

Similar experiences are related by Farrell (7), Campbell (11) and others (12).

Lime based sludges have invariably superior dewatering characteristics than metal salt based sludges. This is well documented in the literature (7, 13, 14).

4 SLUDGE UTILIZATION/DISPOSAL

Sludge Characteristics

One of the factors impacting on potential sludge utilization schemes is that of sludge characteristics.

Sludge characteristics are modulated not only by the type of waste treatment processes employed but are a function of the constituent inputs to municipal sewerage systems. More specifically, industrial discharges to municipal sewers may contain heavy metals, nitrogenous compounds, phosphates, a diversity of complex organic compounds, etc. In biological and physical/chemical treatment systems, most of these compounds are complexed, broken down and/or sorbed by sludge flocs. In most instances, additional chemicals such as lime and/or iron salts are added to enhance the dewatering characteristics of the sludge. Furthermore as phosphorus removal is practiced, sludges not only contain appreciable amounts of phosphorus, the precipitant used to complex the phosphorus, but higher metal concentrations.

In sludge application to land, a number of factors must be considered. For instance, the heavy metal concentration in the sludge will dictate the total amount of sludge that can safely be applied over the lifetime of a site. The "total" metals are generally considered as an indicator of the likely ultimate effect and is used by many to calculate sludge application rates. If an "immediate" effect needs to be ascertained, then this is represented by the "available" fraction of the metal(s) as determined using suitable reagents for extraction. It is important to recognize the potential cumulative and thus, long term effect of metal addition to soil in excess of the small amounts taken up by plants and that lost due to leaching.

The nitrogen content of a sludge will dictate the annual application rate and should be consistent with use of nitrogen by agronomic crops. This will reduce the potential for nitrate pollution of groundwater.

Sludge as a Source of Pollutants

As noted earlier, the major problem in the area of sludge utilization on land concerns the content of potentially toxic substances. However, the level and nature of the substance(s) will also dictate the choice of land utilization alternative. More specifically, sludge may be used as a soil builder and/or organic fertilizer, for land reclamation or application to agricultural land.

Sludge from wastewater treatment facilities practicing chemical phosphorus removal contains almost all of the metals which are discharged into sewers. In the case of heavy metals occurrence in Ontario sludges, Table 7 summarizes this information from 40 Ontario water pollution control plants (10 primary, 30 secondary).

TABLE 7. ONTARIO FLUID SLUDGES - HEAVY METAL CONCENTRATIONS [6]

Component	Primary Plants* Anaerobically digested Sludges			Secondary Plants ** Anaerobically and Aerobically Digested Sludges		
	Range mg/l	Mean*** mg/l	Stand. Dev. ± mg/l	Range mg/l	Mean mg/l	Stand. Dev. ± mg/l
Zinc	2.8 - 130	74.3	48.3	4 - 225	55.5	57.4
Copper	4.6 - 150	54.5	46.9	7 - 148	34.8	30.2
Nickel	0.7 - 15	4.4	4.7	0.26 - 16.8	6.5	14.9
Chromium	2 - 68	16.5	20.9	2 - 430	41.6	51.7
Lead	11 - 86	40.9	30.3	3.7 - 60	21.8	25.1
Cadmium	0.2 - 2.6	0.7	0.7	0.1 - 8.7	1.4	2.0
Cobalt	<0.6 - 1.4	1.0	0.3	0.3 - 3.6	0.8	0.8

*No. of Plants = 10

**No. of Plants = 30

***Arithmetic Mean

The concentrations of heavy metals in digested sludges from primary and secondary plants are similar except for chromium which is three times lower in primary digested sludges.

In 1975, approximately 34% (53 000 tons dry weight) of the sludge produced in Ontario was applied to agricultural land. This resulted in annual, heavy metal loadings as shown in Table 8.

TABLE 8. ANNUAL HEAVY METAL LOADINGS TO SLUDGED ONTARIO SOILS (1975 ESTIMATE) 6

Metal	Average mg/kg soil	Average kg/ha *	Total tonnes
Zn	3.4	6.9	76.7
Cu	2.2	4.4	48.9
Ni	0.4	0.8	8.5
Cr	2.4	4.8	53.1
Pb	1.4	2.8	31.4
Cd	0.1	0.2	1.8
Co	0.05	0.1	1.1

* multiply by 5.5 = lbs/acre

Sludge as a Fertilizer

Major plant nutrients, nitrogen, phosphorus and potassium are contained in sewage sludge. Typical concentrations are 3% nitrogen, 2.5% phosphorus and 0.5% potassium on a dry weight basis. The nutrients in sludge are at a level of 1/5 of the usual chemical fertilizers.

Sludge quality data from the Ontario sludge survey (6) were obtained for 43 water pollution control plants (10 primary, 33 secondary). Forty of these plants disposed of sludge in fluid form, two disposed of sludge cake and one disposed of composted sludge. Table 9 summarizes data on TS, VS, Ammonia

Nitrogen, Total Kjeldahl Nitrogen (TKN) and Phosphorus for digested sludges at primary and secondary plants.

TABLE 9. ONTARIO FLUID SLUDGES - NUTRIENT CHARACTERISTICS 6

Constituent	Primary Plants* Anaerobically Digested Sludge			Secondary Plants** Anaerobically Digested Sludge			Secondary Plants*** Aerobically Digested or Waste Activated Sludge		
	Range	Mean	Stand. Dev. $\pm\sigma$	Range	Mean	Stand. Dev. $\pm\sigma$	Range	Mean	Stand. Dev. $\pm\sigma$
Total Solids - TS%	2.8 - 12.5	8.8	2.9	2.0 - 12	4.1	1.8	2.2 - 4.5	2.75	0.95
Volatile Solids - VS%	24 - 61	43.4	10.5	36 - 70	51	8.5	41 - 69	55.8	9.9
Ammonia Nitrogen - N mg/l	100 - 590	326	78	250 - 1200	628	245	20 - 180	110	24.8
Total Kjeldahl Nitrogen - N mg/l	950 - 2900	1736	913	1600 - 3000	2114	495	650 - 2300	1358	576
Total Phosphorus - P mg/l	240 - 2600	713	399	390 - 2900	975	603	440 - 1200	730	303

* No. of Plants = 10
** No. of Plants = 25
***No. of Plants = 8

In applying 296 x 10^6 gallons (53 000 dry wt tons) of digested sludge to agricultural land in Ontario, the amount of nutrients applied during one year are summarized in Table 10. The data also indicate a relationship between TKN and total solids for anaerobically digested sludge from 23 secondary plants practicing metal salt addition for P removal (Figure 19). This relationship can be used to calculate the TKN loading to farmland as follows:

$$\% \text{ TKN} = 16.6 \ (\% \text{ TS})^{-0.799} \qquad (6)$$

The survey data showed an ammonia nitrogen to TKN ratio varying between 12 and 57% (average 30%). The ammonia nitrogen loading to farmland can thus be approximated by using this average value.

TABLE 10. SLUDGE NUTRIENTS APPLIED ANNUALLY TO ONTARIO FARMLAND (ESTIMATE)[6]

Constituent	Amount
	tonnes/yr
TKN	2 800
NH_4-N	810
Total P as P	1 280
K	270
TS	53 000
VS	30 000

to produce 1 ton of nitrogen fertilizer 750 gal fuel

1500 → incinerate 1 ton

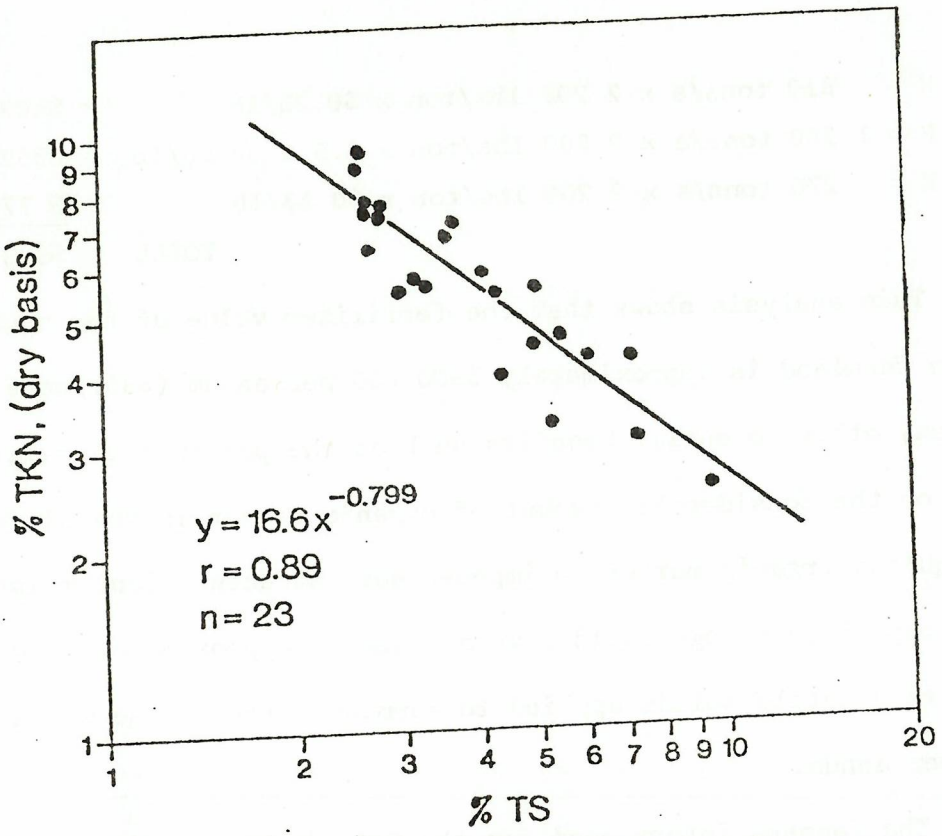


FIGURE 19. TOTAL KJELDAHL NITROGEN VS. TOTAL SOLIDS IN ANAEROBICALLY DIGESTED SLUDGE AT STANDARD C.A.S. PLANTS [6]

By making a number of assumptions it is possible to estimate the potential nutrient value of sludge when applied to farmland in Ontario:

1. the 'available' nitrogen in fluid sludge is equal to the soluble nitrogen (NH_4-N) (a conservative estimate),
2. only one-half of the total phosphorus in liquid sludge is potentially plant available (15) (a conservative estimate),
3. the 'available' potassium in fluid sludge is equal to the total potassium,
4. based on commercial fertilizer prices (October, 1978) the prices for nitrogen, phosphorus and potassium are 23¢, 25¢ and 13¢ per lb respectively.

Using the amounts of nutrients applied to farmland as shown in Table 10 and the aforenoted assumptions, the sludge fertilizer value can be calculated as follows:

NH_4-N =	810 tons/a x 2 200 lbs/ton x \$0.23/lb	≈ \$409 900
P =	1 280 tons/a x 2 200 lbs/ton x 0.5 x \$0.25/lb	≈ \$352 000
K =	270 tons/a x 2 200 lbs/ton x \$0.13/lb	≈ \$ 77 200
	TOTAL	≈ \$839 100

This analysis shows that the fertilizer value of the sludge now applied to farmland is approximately \$900 000 per annum (\$30./acre). This excludes any other potential benefits such as the presence of calcium, magnesium or the considerable amount of organic matter in the sludge. If the farmer required organic matter to improve soil structure (and moisture retention capacity) sludge could have a value of approximately \$20./ton. Based on the volatile solids applied to farmland, the sludge would be worth \$600 000 per annum.

The farmers interviewed for the Ontario survey (6) attempted to quantify yield increases due to sludge application. The average increase in hay yield was estimated at 8 tonnes/ha and in corn yield at 1 tonne/ha. Benefits resulting from cattle weight gain were also noted. A reasonable estimate of the benefits of sludge use on agricultural land in Ontario lies somewhere between \$2 000 000 and \$3 000 000 per annum (16). At this time, the farmers receive sludge free of charge with the transportation costs charged against disposal costs, borne by the municipalities. In 1975, sludge haulage to farmland costs were approximately \$2 250 000.

To date, no negative effects on crop yields were reported by farmers applying sludge for periods in excess of five years. However, long term studies are required to assess whether heavy metals will have negative effects on plants, soil or leachate.

Field monitoring of soil, plants and leachate quality at selected sites where sludges containing high concentrations of heavy metals have been applied for extended periods, would be desirable. A preliminary study in this regard is in progress at the Wastewater Technology Centre (17).

Continuing Studies

Investigations concerning sludge/soil interactions at laboratory greenhouse, field trial and lysimeter scale have been in progress since 1973 at the University of Guelph (18-22) and the Wastewater Technology Centre (23-26). Some of the more significant conclusions from these studies are:

- Sewage sludges supplied nitrogen and phosphorus for crop production but were low in potassium. Sludges produced crop yields at least as high as were obtained with chemical fertilizers.
- Sewage sludge application did not result in marked increases in runoff of nutrients, heavy metals or bacteria on 2% and 6% slopes except when heavy rain occurred immediately after sludge application.
- Soil salinity was not a problem in the field under Ontario conditions. It might pose a problem in less humid areas. Boron levels in some sludges tested would also be expected to pose a problem in arid regions.
- Re-application of the sludges between crops did not lead to increased metal concentrations in the plant materials.
- Large amounts of metals were added to soils in some sludges and their removal by crop uptake or leaching was very limited.
- The organic nitrogen in sludges was mineralized gradually and the mineralization rate varied from one sludge to another. As with other sources of nitrogen, applications in excess of crop requirements lead to high levels of nitrate in the soil solution.
- The average NH_4^+-N content of sludges studied was 1.3% on a dry weight basis or 27% of the total nitrogen. In two experiments 40% and 48% of the NH_4^+-N was lost by volatilization from sewage sludge applied to the soil surface. This loss occurred in five and eight days, respectively.

- Salmonella were isolated from five of 207 sludge samples tested. If vegetables are not grown and animals not grazed immediately following sludge application and if reasonable care is exercised in spreading, sludge does not pose a serious health hazard.
- At least twice as much nitrogen must be applied in fluid sludge as in commercial fertilizer to obtain equivalent yields.
- The Cd, Cr, Cu, Ni, Pb and Zn concentrations in orchard grass and wheat plant materials have not exceeded suggested maximum "tolerance" or "toxic" levels.
- The maximum concentrations of Cd, Cr, Cu, Ni, Pb and Zn in leachates have not exceeded drinking water standards.
- Soluble P in the leachates from fluid sludge treatments ranged as high as 10 mg/L during summer 1977. Soluble P from the air-dried sludge treatments never exceeded 2 mg/L.
- Total organic carbon in leachates from both the fluid and air-dried sludge experiments were greater than 50 mg/L in 1976 at the highest sludge loading rates. Static bioassay toxicity tests using Daphnia showed no toxicity in these leachate samples.

Ontario Guidelines for Sludge Utilization on Agricultural Land

The amount of NH_4-N applied to Ontario farmland in 1975 was 810 tonnes/yr (Table 10). Combining this information with the data on heavy metal application to farmland (Table 8) allows for an assessment as to whether or not Ontario sludge is generally suitable for land application if the criteria of the Provisional Guidelines for Sewage Sludge Utilization on Agricultural Land (27) are applied. This assessment is summarized in Table 11 and shows that the sludge is generally suitable for agricultural land application. The exception is the cadmium content. Sources of high cadmium content sludges are few and isolated. It is important to stress that average

values are extremely misleading and that it is imperative that each sludge source be characterized separately in order to determine its limits of suitability for application to farmland.

TABLE 11. SUITABILITY OF ONTARIO SLUDGE FOR UTILIZATION ON FARMLAND[6]

Constituent[1]		NH_4-N:Heavy Metal		Suitability
NH_4-N	Heavy Metal	Actual Ratio	Minimum Required Ratio[2]	
810				
	Zn 76.7	11	4	Yes
	Cu 48.9	17	10	Yes
	Ni 8.5	95	40	Yes
	Cr 53.1	15	6	Yes
	Pb 31.4	26	10	Yes
	Cd 1.8	450	500	No
	Co 1.1	736	50	Yes

[1] in tons (metric)/yr (Tables 8 and 10)

[2] Provisional Guidelines for Sewage Sludge Utilization on Agricultural Land[27]

NH_4-N + NO_3-N

5 CLOSING REMARKS

The information presented represents a summary of Canada's experience in the Province of Ontario with increased sludge production due to chemical removal of phosphorus to 1.0 mg/L total phosphorus when using metal salts.

Ontario's current sludge management strategy consists of applying the most cost-effective and environmentally acceptable solution. Sludge utilization for its nutrient value on agricultural land is one such management strategy followed by an increasing number of municipalities who, as well as the farmers, are concerned about potentially long-term harmful impacts on soils due to heavy metal addition.

While technological solutions to phosphorus point source control to 0.1 mg/L are available, the impact on sludge quantities generated, handling and disposal still remains to be more closely defined. Only with information on relative costs between alternatives to achieve these goals can an effective point source phosphorus control management strategy be proposed.

Computer simulation is one approach to assess potential management strategies. It may well turn out that point source control to levels substantially lower than currently practiced will cause more problems elsewhere. The current activities of the IJC's Phosphorus Management Strategies Task Force address this subject.

REFERENCES

1. Jank, B.E., and P.H.M. Guo, "Biological Treatment of Meat and Poultry Wastewater" presented at Meat Technology Transfer Seminar on Meat and Poultry Industry Regulations and Guidelines, February, 1978.

2. Kormanik, R.A., "Estimating Solids Production for Sludge Handling", Water and Sewage Works, December, 1972.

3. Campbell, H.W., R.J. Rush and R. Tew, "Sludge Dewatering Design Manual", COA Research Report No. 72, January, 1978.

4. Sutton, P.M., K.L. Murphy and B.E. Jank, "Nitrification Systems with Integrated Phosphorus Precipitation", presented at PCAO Conference, Toronto, April, 1977 and B.C. Water and Waste Association Conference.

5. Stepko, W.E., and D.T. Vachon, "Phosphorus Removal Demonstration Studies Using Lime, Alum and Ferric Chloride at C.F.B. Borden", Environmental Protection Service, EPS 4-WP-78-2, Water Pollution Control Directorate, Ottawa, February, 1978.

6. Antonic, M., M.F. Hamoda, D.B. Cohen and N.W. Schmidtke, "A Survey of Ontario Sludge Disposal Practices", Project No. 74-3-19, COA Research Report (in press).

7. Farrell, J.B., "Design Information on Dewatering Properties of Wastewater Sludges", Sludge Handling and Disposal Seminar, COA Conference Proceedings No. 2, Toronto, 1974.

8. Brouzes, R.J.P., "The Use of Lime in the Treatment of Municipal Wastewaters", COA Research Report No. 21, Ottawa.

9. Prested, B.P., E.E. Shannon and R.J. Rush, "Development of Prediction Models for Chemical Phosphorus Removal, Volume II", COA Research Report 78, June, 1978.

10. Drynan, W.R., "Relative Costs of Achieving Various Levels of Phosphorus Control at Municipal Wastewater Treatment Plants in the Great Lakes Basin", Technical Report to the International Reference Group on Great Lakes Pollution from Land Use Activities of the International Joint Commission, July, 1978.

11. Campbell, H.W., R. Tew and B.P. Le Clair, "Some Aspects of Chemical Sludge Thickening and Dewatering", presented at the Alternatives for Nutrient Control Seminar, Kelowna (1975).

12. Moss, W.H., R.E. Schade, S.J. Sebesta, K.A. Scheutzew, P.V. Beck and D.B. Gerson, "Full-scale Use of Physical/Chemical Treatment of Domestic Wastewater at Rocky River, Ohio", Jour. Water Poll. Control Fed., 49, November, 1977.

13. Stickney, R., and B.P. Le Clair, "The Use of Physicochemical Sludge Characteristics and Bench Dewatering Tests in Predicting the Efficiency of Thickening and Dewatering Processes", Sludge Handling and Disposal Seminar, COA Conference Proceedings No. 2, Toronto, 1974.

14. EPA, "Process Design Manual for Sludge Treatment and Disposal", U.S. EPA Technology Transfer Report, EPA 625/1-74-006, October, 1974.

15. U.K. Department of the Environment, National Water Council, "Report of the Working Party on the Disposal of Sewage Sludge to Land", Standing Technical Committee Report Number 5, London, England.

16. Schmidtke, N.W., and D.B. Cohen, "Municipal Sludge Disposal on Land, A Down-to-Earth Solution", presented at the Western Canada Water and Sewage Conference, Edmonton, Alberta, September 28-30, 1977.

17. Monteith, H., D.N. Bryant and M.D. Webber, "Assessment of PCB's and Heavy Metals at Selected Disposal Sites in Ontario", Project No. 76-3-26, COA Research Report (in preparation).

18. Bates, T.E., E.G. Beauchamp, R.A. Johnston, J.W. Ketcheson, R. Protz and Y.K. Soon, "Land Disposal of Sewage Sludge - Volume 1", COA Research Report No. 16, Ottawa, 1975.

19. Bates, T.E., E.G. Beauchamp, R.A. Johnston, J.W. Ketcheson, R. Protz and Y.K. Soon, "Land Disposal of Sewage Sludge - Volume II", COA Research Report No. 24, Ottawa, 1975.

20. Bates, T.E., E.G. Beauchamp, R.A. Johnston, J.W. Ketcheson, R. Protz and Y.K. Soon, "Land Disposal of Sewage Sludge - Volume III", COA Research Report No. 35, Ottawa, 1976.

21. Bates, T.E., E.G. Beauchamp, R.A. Johnston, J.W. Ketcheson, R. Protz and Y.K. Soon, "Land Disposal of Sewage Sludge - Volume IV", COA Research Report No. 60, Ottawa, 1975.

22. Bates, T.E., E.G. Beauchamp, R.A. Johnston, J.W. Ketcheson, R. Protz and Y.K. Soon, "Land Disposal of Sewage Sludge - Volume V", COA Research Report No. 73, Ottawa, 1978.

23. Chawla, V.K., J.P. Stephenson and D. Liu, "Biochemical Characteristics of Digested Chemical Sewage Sludges", COA Conference Proceedings No. 2, 1974.

24. Chawla, V.K., D.N. Bryant, D. Liu and D.B. Cohen, "Chemical Sewage Sludge Disposal on Land - (Lysimeter Studies) - Volume 1", COA Research Report No. 67, Ottawa, 1977.

25. Cohen, D.B., M.D. Webber and D.N. Bryant, "Land Application of Chemical Sewage Sludge "Lysimeter Studies", Sludge Utilization and Disposal Seminar, Toronto, Ontario, 1978.

26. Cohen, D.B. and D.N. Bryant, "Chemical Sewage Sludge Disposal on Land - (Lysimeter Studies) - Volume II", COA Research Report No. 79, Ottawa, 1978.

27. OMAF (Ontario Ministry of Agriculture and Food) and OMOE (Ontario Ministry of the Environment) Ad Hoc Joint Committee to prepare "Guidelines for Sewage Sludge Utilization on Agricultural Lands", April 1978.

PHOSPHORUS REMOVAL DESIGN PITFALLS

AND

HOW TO AVOID THEM

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TABLE OF CONTENTS

		PAGE
I	INTRODUCTION	1
II	PLANNING.....	2
III	CONSTRUCTION.....	6
	A) GENERAL.....	6
	B) STORAGE.....	7
	C) CHEMICAL FEED SYSTEMS.....	9
	D) CHEMICAL FEED LINES.....	11
IV	OPERATION.....	13
	BIBLIOGRAPHY.....	15

INTRODUCTION

The topic can be broken down into three subtopics by considering the following areas:

PLANNING - assess your needs
- take an overview of what is available to do the job you want done and decide between several alternatives.

CONSTRUCTION - designing and building the physical facility to perform the required task.

OPERATION - making the facility perform as required.

In each of these three areas serious errors can be made; any one of which will make the final product undesirable.

PLANNING

Considerable emphasis must be placed on the predesign work carried out before implementing phosphorus removal at any pollution control facility. No process for phosphorus removal is universally applicable to all situations. Up until recently only one reliable process type was available for phosphorus removal, namely, chemical precipitation. More recently considerable credibility has been won by various processes of biological phosphorus and nitrogen removal. In most cases the biological processes involve subjecting the wastewater to alternating aerobic, anoxic and anearobic conditions with or without sludge recycle at each stage.

At this time,especially when considering plant expansions or new plant construction, biological means of nutrient removal should be considered. This however cannot be accomplished without carrying out a detail pilot scale project to generate design information. Without such pilot work a significant risk of process failure or high cost factors is inevitable. Even with the more conventional chemical precipitation process such pre-design work is essential.

This paper will not deal with biological systems because of the relative newness of the processes and the lack of specific universally applicable design information. The phosphorus removal system using chemical precipitation has become well established and will be the prime subject of this paper.

The performance of jar tests to evaluate the effectiveness of various chemical precipitants in phosphorus removal has been well documented by the Ontario Ministry of the Environment.(1)

Three specific points should be stressed:

1. All available precipitants should be used in the jar testing program. This will ensure that the most cost effective precipitant is evaluated. A partial list of sources of metal ions for phosporus precipitant is shown in Table I.

TABLE I

SOURCES OF METAL ION FOR PHOSPHORUS REMOVAL

Alum (Aluminum sulphate)

Sodium aluminate

Anodizing wastes (Aluminum ion source)

Ferric chloride

Ferrous chloride

Ferrous sulphate

Waste pickle liquor (Iron ion source)

Lime

Waste carbide lime

The use of waste products and some of their advantages and drawbacks is discussed later.

2. In many plants there is a significant predictable change in wastewater characteristics depending upon the time of year. For example, seasonal food processing industries which discharge to the sewer system can alter the wastewater sufficiently to preclude the use of a specific precipitant for a large percentage of the year. If such a situation does exist it is essential to carry out the jar testing program or any other pilot program during both time periods. It is not necessary to consider

the dilution affects of the spring runoff as this does not significantly affect the precipitation characteristics of the sewage.

3. The use of polyelectrolytes to enhance clarification and improve settling characteristics may also be investigated at this time if the hydraulic load on the final clarifiers frequently exceeds 35 $m^3/m^2/d$ (800 gpd/ft^2). With hydraulic loads less than this value the use of polymers is not warranted.

CONSTRUCTION

GENERAL:

In any chemical precipitation process design there are three basic components; the precipitant storage tanks, chemical feed pumps, and precipitant distribution system. All three of these components should be compatible with all types of available precipitants. This does not apply to lime feed systems as they are unique in that they are designed for a single specific precipitant. The use of lime has not been widespread since its capital and operating costs are high and the system is not compatible with the existing activated sludge systems and high quality effluent.

In the case of liquid precipitants the most corrosive chemical is usually ferric chloride. The only metal suitable for handling ferric chloride is titanium. The use of titanium, of course, would be prohibitavely expensive. There are, however, several plastics available in tanks, pumps and pipe systems. These have proven satisfactory.

Aluminum salts can be safely handled in stainless steel systems. The use of stainless steel is not recommended, since, should the relative price between alum and iron salts change in the future, a stainless steel system would not be able to handle the iron salts.

The control and cleanup of even minor spills should be considered in the design. Spill aprons, washdown lines and appropriate floor drains, sump pumps, etc. are essential design components.

Large truck access to the chemical unloading area is a frequently forgotten aspect of phosphorus removal design.

STORAGE

In most plants with capacities greater than 0.9 M^3/d (0.2 MGD) it is advisable to provide bulk storage of at least 5000 gal. of the precipitant. Heated tanks are needed in Canadian climates to prevent desolubilization of the chemicals. Feeding difficulties may begin to be experienced at temperatures of 7^oC as alum viscosity increases to the point of making

accurate pumping difficult. Temperature control can be attained by installing the tanks inside heated buildings or by heat tracing and insulating outside tanks. The second option has proven to be satisfactory when adding phosphorus removal to existing plants.

Care must be taken to account for thermal expansion of the tank as several of the precipitants may arrive at the plant heated to 49^{o}C (120^{o}F). The materials of construction should also be suitable for this temperature regime.

The provision of tank access ports or cleanouts, exterior venting, raised drawoffs, and volume indicators are good design features.

CHEMICAL FEED SYSTEMS

There are many manufactures of good chemical feed equipment in North America. Care should be taken to ensure the availability of replacement parts and cleanability of the feed heads when considering alternatives. Especially when feeding waste products containing sludge or solid particles,some provision for inline filtration to prevent frequent feed system blockages would be advisable.

At all times a standby pump should be specified. In large plants (10 MGD^{+}) the installation of three feed pumps with two in use is recommended. Smaller plants can be adequately served with a two pump system. Flow matching of pump feed rates is expensive for 1 MGD plant but does provide a uniformity of feed rate that improves process performance and eliminates chemical wastage if properly adjusted. In plants greater than 2 MGD economic considerations make the installation of flow matching equipment advisable. The pumps should be sized to provide sufficient feed for the highest volume precipitant.

When the chemical storage tank's maximum liquid level is

above the elevation of the pump discharge point, anti siphon provisions are essential.

The use of visual flow rate indicating devices has been tried in a number of installations. To date none of these attempts have proven to be satisfactory. It is, however, a desirable quality to be able to visually observe and/or record the chemical feed rate.

CHEMICAL FEED LINES

The chemical distribution system should have a great deal of flexibility and control. At least three points of precipitant addition should be available; entrance to the primary clarifiers, mid point of the aeration tanks, and between the aeration tanks and the secondary clarifier. The purpose of adding chemicals prior to the primary clarifier would be to help reduce an organic overloading situation by improving the BOD removal of the primaries. The mid point of the aeration tank is the appropriate point to add any sources of ferrous ion as it must be oxidized to ferric ion prior to sedimentation. The increased load on the aeration system to provide oxidation of the ferrous ion to ferric ion is so small it can be ignored for all practical purposes. Chemicals such as alum or ferric chloride which provide almost instantaneous floc formation and phosphorus precipitation are best added at the discharge of the aeration tank.

It is essential for efficient chemical usage and effective phosphorus precipitation to provide good mixing at the point of chemical addition. Mixing can

frequently be found within the existing plant at Parshall flumes, overflow weirs, etc. If no suitably mixed point can be found for precipitant addition supplementary mixing must be provided.

A chemical distribution board at the feed pump location is helpful for process control.

The feed lines should be buried below the frost line or be heat traced where exposed to freezing temperatures. The discharge points should be above the water level and accessable to plant personnel in order to provide a convenient confirmation of chemical feed rates.

Particularly in large systems some means of surge control should be provided adjacent to the feed equipment to prevent premature line failure. Expansion of lines due to large temperature changes should also be considered.

OPERATION

The day to day operation of a properly designed phosphorus removal system is limited to monitoring the chemical consumption and effluent phosphorus levels daily. If the soluble phosphorus level drops below 50% of the required total phosphorus effluent standard more chemical than required is being added. A daily calculation of the precipitant dosage serves as a check on the feed system. By comparing the dosage and the effluent level a measure of the process effectiveness can be determined. The most frequent causes for process failure is under or over dosing of the precipitant. Both cases result in higher than desired total phosphorus levels in the effluent. They can be differentiated by checking the effluent soluble phosphorus level.

The use of waste products frequently complicates operational practises as frequent partial pump blockages, periodic storage tank cleanings and possible toxic metal contamination problems are possible. These additional operating problems must be weighed against the significantly reduced precipitant cost.

Most other operational problems, and a host of them do

exist, relate to poorly designed systems and the attempts of the operational staff to live with or circumvent the specific design faults. These problems are generally site specific and should be rectified on a site specific basis.

BIBLIOGRAPHY

Van Fleet, G.L., Guidelines for Conducting Treatability for Phosphorus Removal at Wastewater Treatment Plants, Ontario Ministry of the Environment, Sanitary Engineering Branch, April, 1972.

COST OF NUTRIENT REMOVAL

by

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Vice President, Technical Services
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Nutrient removal process selection necessarily is based upon effluent requirements and process reliability. Where alternate processes are technically feasible, then the ultimate selection would be based on an economic evaluation. It must be emphasized that process costs are highly site specific and are influenced by the actual waste characteristics. Successful cost evaluation must recognize these constraints.

Phosphorus Removal

Considerable information on the methodology of phosphorus removal and associated costs has been generated as a result of the Canada-Ontario Agreement on Great Lakes Water Quality. Iron salts, aluminum salts and lime are all proven precipitants and are capable of attaining an effluent requirement of 1.0 mg/L total phosphorus. Selection of the appropriate chemical and point of addition would be based upon compatability with the existing facilities, anticipated operating difficulties, and operating and capital costs.

Capital costs associated with phosphorus removal are relatively independent of precipitant or point of addition. Typical phosphorus removal schemes were portrayed schematically by Schmidke and Cohen (1977) and are presented in Figure 1. A typical breakdown of required components and costs for a 1 mgd wastewater treatment plant is presented in Table 1.

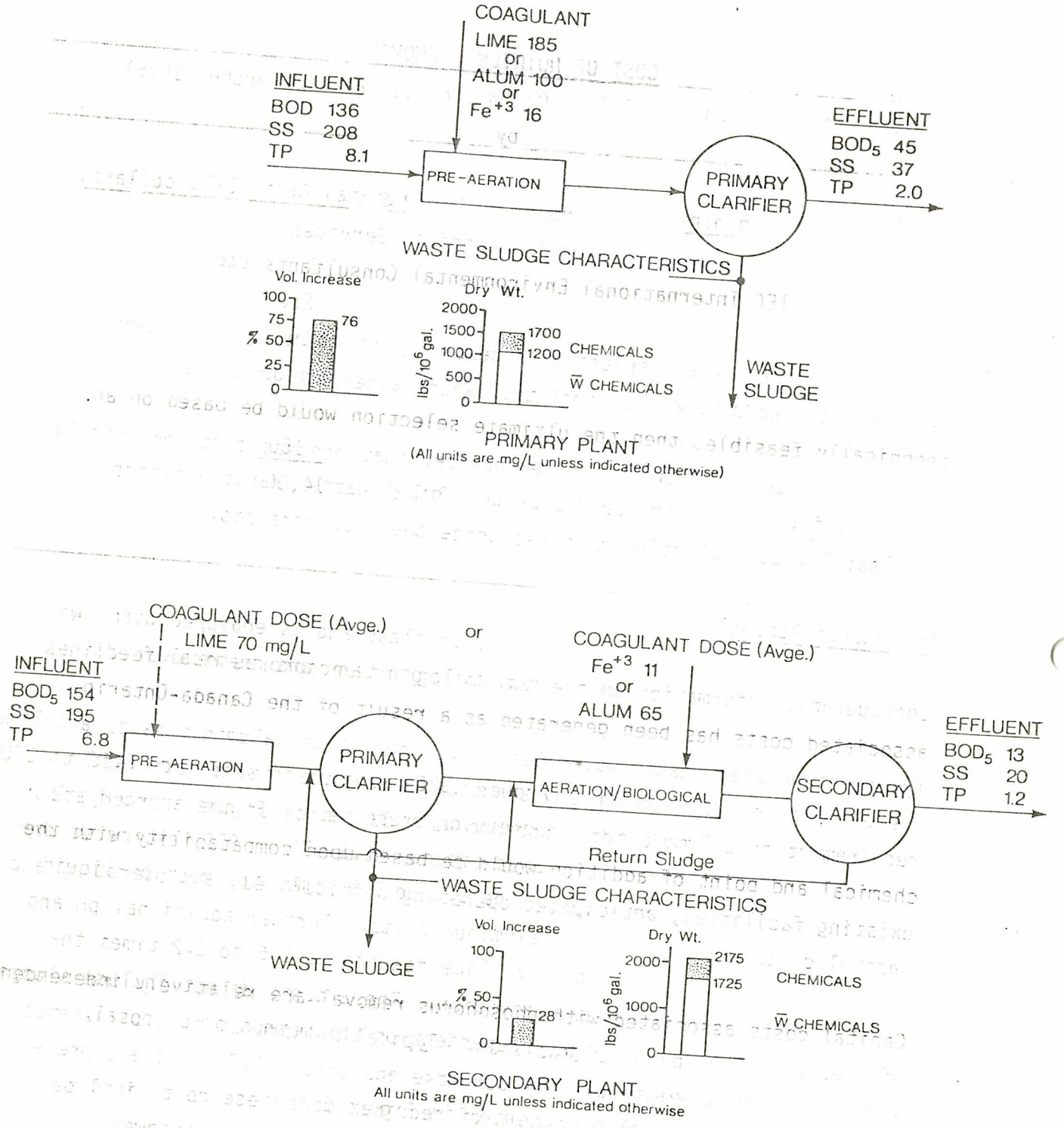


FIGURE 1. TYPICAL PHOSPHORUS REMOVAL PROCESSES (after Schmidtke and Cohen, 1977)

TABLE 1: CHEMICAL FEED EQUIPMENT FOR PHOSPHORUS REMOVAL (Archer 1975)

Component	Capital Cost (1975 dollars)
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Mechanical	1000
Electrical	1000
Structural	1000
Engr. Cont.	1500
TOTAL	14,000

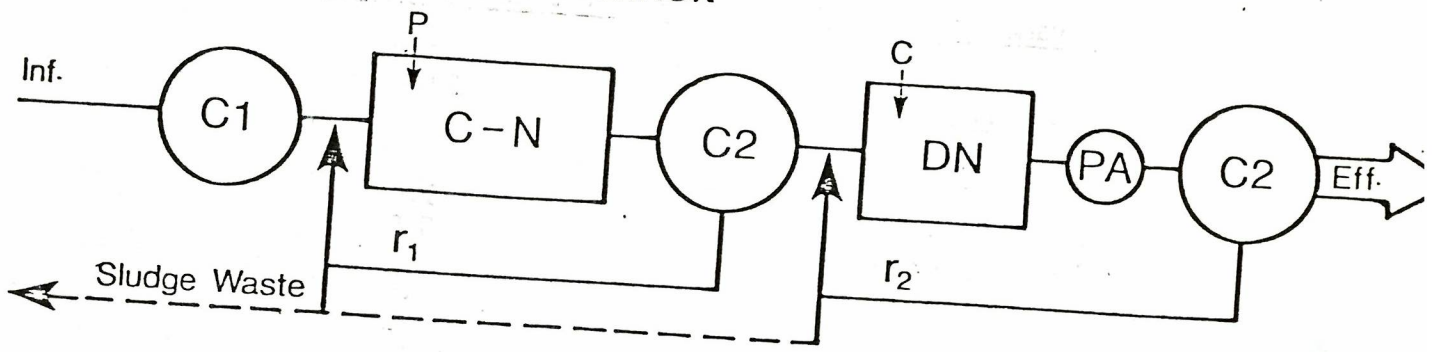
The chemical feed tank is reinforced fibreglass and is equipped with two variable speed chemical metering pumps. Both tank and chemical feedlines are heated.

A survey of 64 Ontario plants (Archer 1975) ranging in size from 0.12 to 180 mgd produced the capital cost regression presented in Figure 2. Only four of these plants used lime as a precipitant. U.S. EPA (1976) estimated dry chemical feed and lime storage to be 1.3 to 1.9 times the cost of liquid feed systems used for iron and aluminum salts. Further additional pH and flow control increased the cost of lime feeding to 1.6 to 2.2 times the capital cost for metallic salt addition. These costs do not include sludge disposal costs which are highly dependent on the method of disposal.

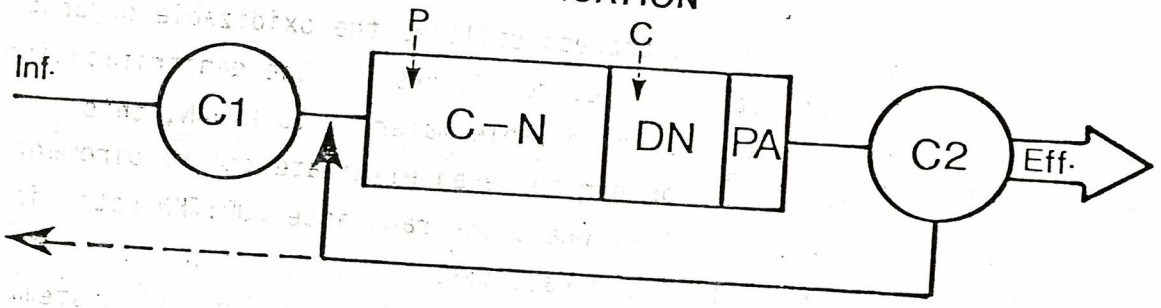
Actual operating costs are presented in Figure 3. These costs include replacement parts, electricity, make-up water and direct manpower maintenance costs, but exclude amortization and apparently associated operating labour costs. Current 1979 chemical costs in Ontario are given in Table 2.

Figure 4
Schematic of Nitrogen Removal Processes

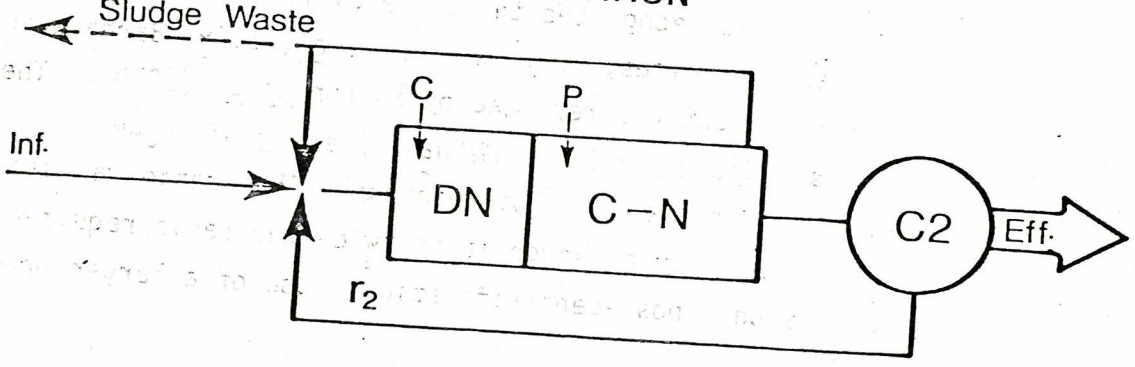
TWO–SLUDGE POST–DENITRIFICATION



ONE–SLUDGE POST–DENITRIFICATION



ONE–SLUDGE PRE–DENITRIFICATION



C1 = Primary Clarifier
C2 = Secondary Clarifier
C–N = Carbon Removal plus Nitrification
DN = Denitrification
Inf = Degritted, screened Wastewater
Eff. = Final Treated Effluent
r = Recycle
PA = Post Aeration
C = External Carbon Source
P = Chemical Addition for P Removal

TABLE 3: EFFLUENT OBJECTIVES, PENTICTON, B.C.

Parameter	Treated Effluent (mg/L)
Total BOD_5	20
Suspended Solids	20
NH_4 - N	1.0
$NO_3 + NO_2$ - N	5.0
Total P	0.6
pH	>6.5

The one-sludge pre-denitrification process utilizes the oxidizable organic substances in the raw wastewater as electron donors for the denitrification process. Depending on the ratio of oxidizable material to NO_3-N, this process configuration can reduce or perhaps even eliminate the requirement for methanol addition. At Penticton, where the raw waste COD:TKN ratio is 10:1, there is insufficient COD in the wastewater to allow complete denitrification and additional methanol was required in the pre-DN system. For this study 14 kg of total raw wastewater COD per kg of NO_3-N equivalent was selected as the required ratio for the internal carbon source for pre-denitrification. It is recognized there are differing views in the literature (Christensen and Harremoes, 1977, Sutton et al, 1979, Barnard and Meiring, 1977). This requirement was reported by Sutton et al (1979). They also established that approximately 6 kg of methanol per kg of NO_3-N equivalent was needed for their pre-DN pilot reactors. This ratio was used in the calculations for this study although it is twice the ratio required for either 1-sludge or 2-sludge post-denitrification. Use of a larger pre-

denitrification reactor, where endogenous respiration would provide the carbon source for denitrification, was rejected as no reliable information was available for low temperatures. Oxygen requirements are less than for the other two processes since most of the oxidizable organics are removed during pre-denitrification before the wastewater enters the aerated portion of the basin.

Capital costs were developed using U.S. Environmental Protection Agency (EPA) cost estimating guidelines (Patterson and Banker, 1971), CH_2M Hill Ltd. cost data and experience, and manufacturer cost information. The EPA guidelines were developed from actual construction costs on a large number of treatment plant projects. Adjustments have been made in the costs to account for local economic conditions.

Annual operation and maintenance costs were determined by estimating labour, electric power and chemical requirements, and equipment replacement and supply costs. Labour requirements were estimated using EPA's staffing guidelines, (CH_2M-Hill, 1973), and information published by Culp et al, (1977). Cost data are presented in January, 1979 dollars.

The capital costs for the basic components were increased 15 percent for yardwork, 20 percent for contingencies, and 20 percent for engineering, legal and administrative items. Yardwork includes miscellaneous piping, plumbing, electrical, instrumentation and control equipment, grading and paving.

O & M estimates were based on unit costs for labour, power and chemicals at Penticton presented in Table 4.

TABLE 4: BASIS FOR OPERATING AND MAINTENANCE COST ESTIMATES

Item	Unit Cost
Labour	$22,000/man-year
Power	$0.01/kilowatt-hour
Lime	$72.70/tonne
Methanol	$0.22/litre
Pickle Liquor	$2.11/100 kilograms
Chlorine	$319/tonne
Polymer	$5.29/kilogram
Equipment Replacement and Supply	0.6% total capital cost/year

Capital, plus operating and maintenance cost estimates for the biological nitrogen removal process are presented in Table 5. The cost for conventional activated sludge treatment and nitrification have also been presented for comparison. Only those components of the treatment plant required for carbon and nitrogen removal are included in the cost comparison. This excludes pretreatment, sludge digestion and disposal, chlorination, and final treated effluent discharge.

Present worth and total annual costs of the five processes in Table 5 are in January, 1979 dollars. Present worth values are based on 20 years of operation and maintenance at a discount rate of 10.5 percent. Capital costs for each process have been amortized at 10.5 percent interest for 20 years and added to the annual O & M cost to establish the total annual cost.

The comparative ratings presented in Table 5 indicate that the total cost of the 2-sludge process is 15 to 20% higher than for the 1-sludge systems. Although the pre-denitrification system has about the same capital cost as

Item	Cost
Labour	$22,000/man-year
Power	
Lime	
Reagents	
Pickle liquor	$2.11/100 kilograms
Chlorine	
Pit area	
Equipment Replacement and Supply	6.0% total capital cost/year

chlorination, and final treated effluent discharge.

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TABLE 5: ECONOMIC COMPARISON OF BIOLOGICAL NITROGEN REMOVAL PROCESSES

	Activated Sludge	Nitrification	Two-Sludge Post-DN	One-Sludge Post-DN	One-Sludge Pre-DN
Present Worth[a]					
Capital	$ 2,618,000	$ 3,268,000	$ 4,802,000	$ 4,180,000	$ 4,217,000
Present worth of annual O&M	1,103,000	1,679,000	3,366,000	2,996,000	2,567,000
Total present worth	$ 3,721,000	$ 4,947,000	$ 8,168,000	$ 7,176,000	$ 6,784,000
Comparative rating	0.55	0.73	1.20	1.06	1.00
Annual Cost					
Amortized capital[a]	$ 318,000	$ 397,000	$ 583,000	$ 508,000	$ 512,000
O&M	134,000	204,000	409,000	364,000	312,000
Total annual cost	$ 452,000	$ 601,000	$ 992,000	$ 872,000	$ 824,000
Comparative rating	0.55	0.73	1.20	1.06	1.00

[a] Based on a 10.5-percent discount/interest rate for 20 years

NOTE: All costs are in January 1979 dollars

TABLE 6: COST COMPARISON OF BIOLOGICAL NITROGEN REMOVAL WITH OTHER METHODS

	Activated Sludge (x $1,000)	Nitrification (x $1,000)	Biological Nitrogen Removal (x $1,000)			Other Methods (x $1,000)		
			Two-Sludge Post-DN	One-Sludge Post-DN	One-Sludge Pre-DN	Clinoptiolite Ion Exchange	Ammonia Stripping	Land Disposal
Capital	$ 7,513	$ 8,600	$10,134	$ 9,512	$ 9,558	$15,980	$12,561	$19,434
Annual O&M	421	485	600	555	503	629	679	162
Present Worth of O&M[a]	3,456	3,992	4,938	4,568	4,140	5,177	5,589	1,333
Total Present Worth	$10,969	$12,592	$15,072	$14,080	$ 13,698	$21,157	$18,150	$19,712[b]
Comparative Rating	0.80	0.92	1.10	1.03	1.00	1.54	1.32	1.44

NOTE: Costs are in January 1979 dollars for a complete treatment facility at Penticton, B.C. with capacity of 1.8×10^4 m^3/d, 4540 kg BOD/d, and average annual flow of 1.55×10^4 m^3/d.

[a] Based on a discount rate of 10.5 percent for 20 years.

[b] Includes pipeline salvage value of $365,000 based on straight-line depreciation and a 40-year pipeline life, land salvage value of $690,000 based on 3-percent compound appreciation in land value for 20 years, and 10.5-percent discount rate.

alone and activated sludge treatment. The curves were prepared from separate cost estimates for a 0.45 x 10^4 m^3/d(1.0 IMGD), a 1.5 x 10^4 m^3/d (3.4 IMGD) and a 4.5 z 10^4 m^3/d (10 IMGD) plant using the wastewater characteristics at Penticton as a design basis. Equipment, power, chemical and labour costs are also based on the Penticton values.

Capital and O & M costs for new facilities presented in Figure 5 include only those portions of the treatment plants associated with carbon removal and nitrogen removal. This excludes headworks, primary clarifiers, sludge processing and disinfection. The capital costs in Figure 5 do not include the 20% contingency included in the costs in Table 5 but do allow for yardwork and engineering.

Figure 6 presents curves for the costs of "add-on" facilities to existing activated sludge treatment plants. The add-on capital costs were prepared by estimating the cost of additional basin volume, aeration, mixing and chemical feed equipment. It was assumed that no major renovations would be necessary to the existing final clarifiers and that they would be able to handle the increased solids loading caused by plant conversion. As before, the capital cost includes yardwork and engineering. A 15% contingency was added due to the uncertainties of predicting costs for renovations to existing plants.

Figure 5 shows that a 10 fold increase in plant size gives a 5 fold increase in capital cost for an activated sludge system and a 6 fold increase for a pre-denitrification system. The difference is caused by the relative difference in contribution to capital cost of the mechanical versus the structural components. For activated sludge systems, mechanical components represent a significantly higher percentage of total capital costs in comparison to pre-denitrification which requires more basin volume. Unit costs for mechanical components increase with plant capacity more slowly than structural components. Therefore, the higher percentage of mechanical components in an activated sludge plant is reflected in a lower rate of increase in total capital cost with increasing plant size.

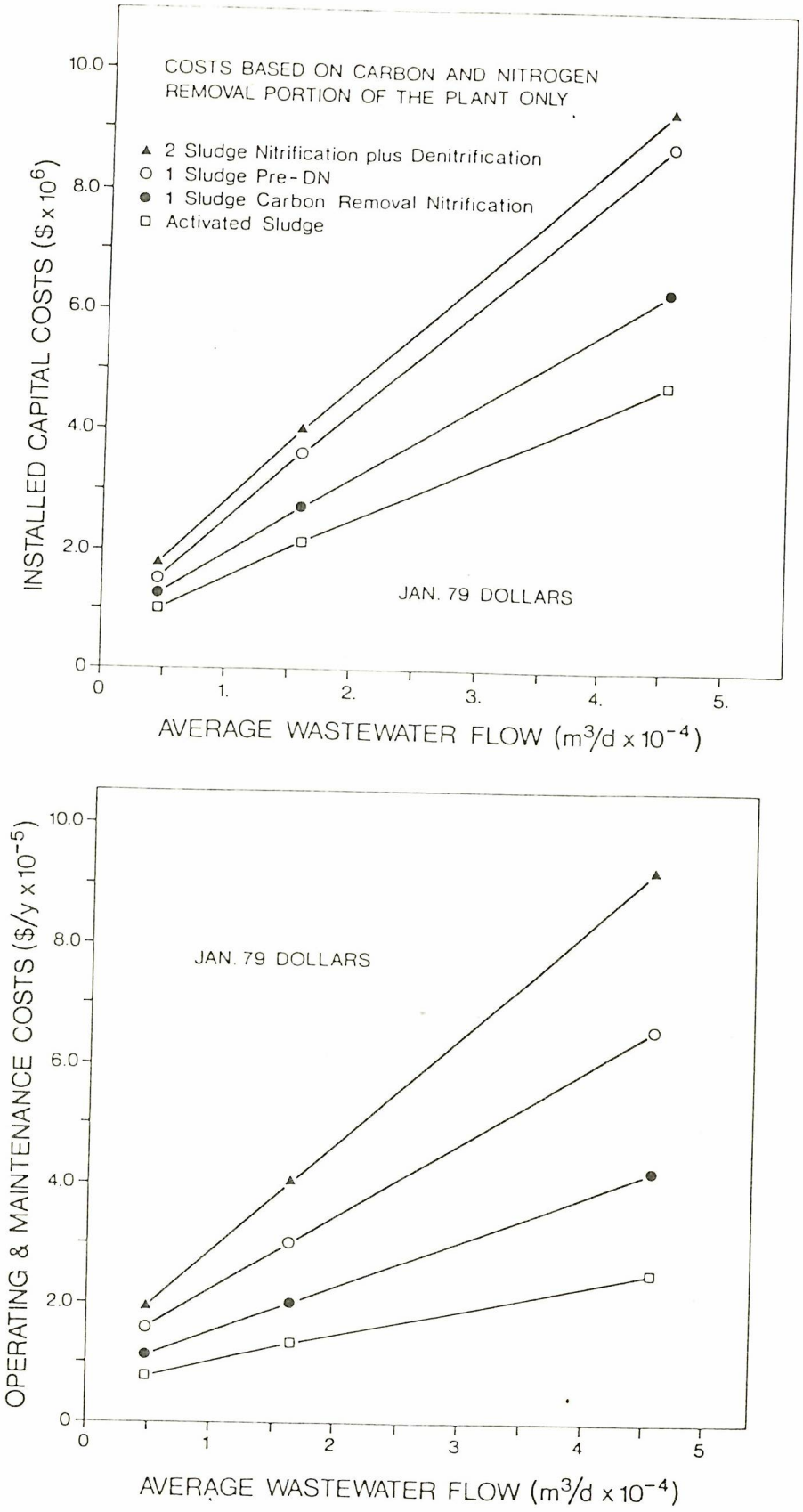
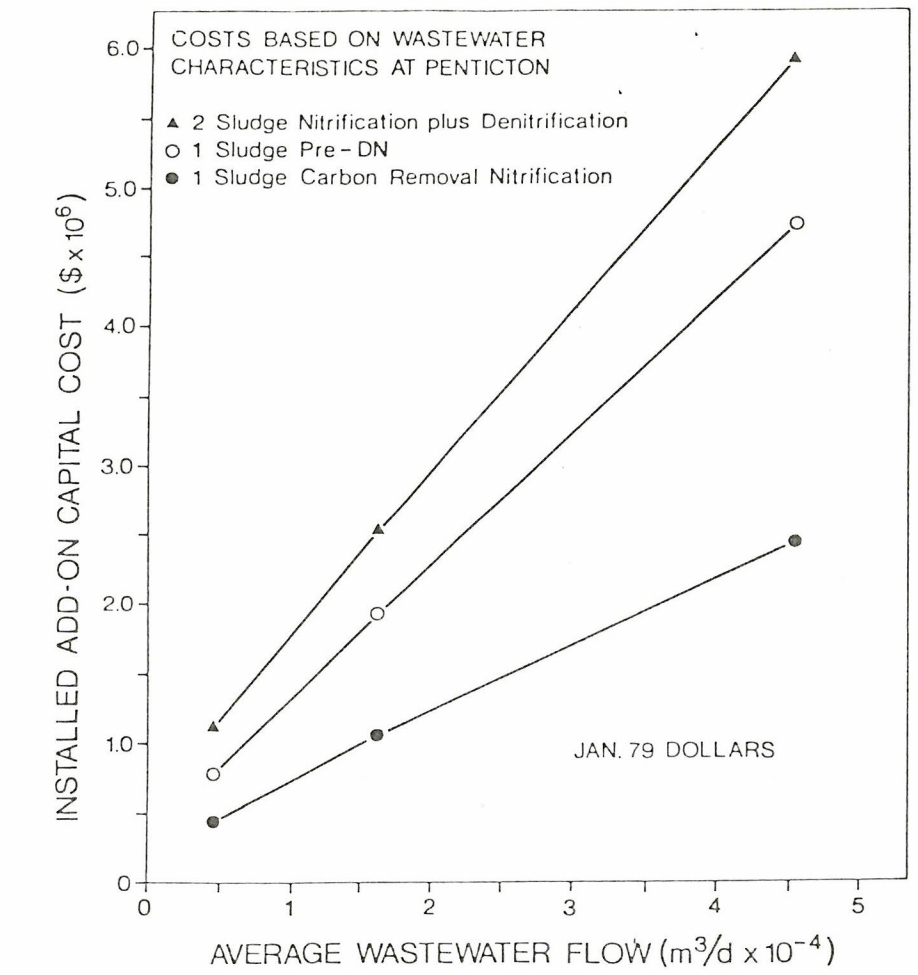


FIGURE 5. COSTS FOR CARBON AND NITROGEN REMOVAL PROCESSES



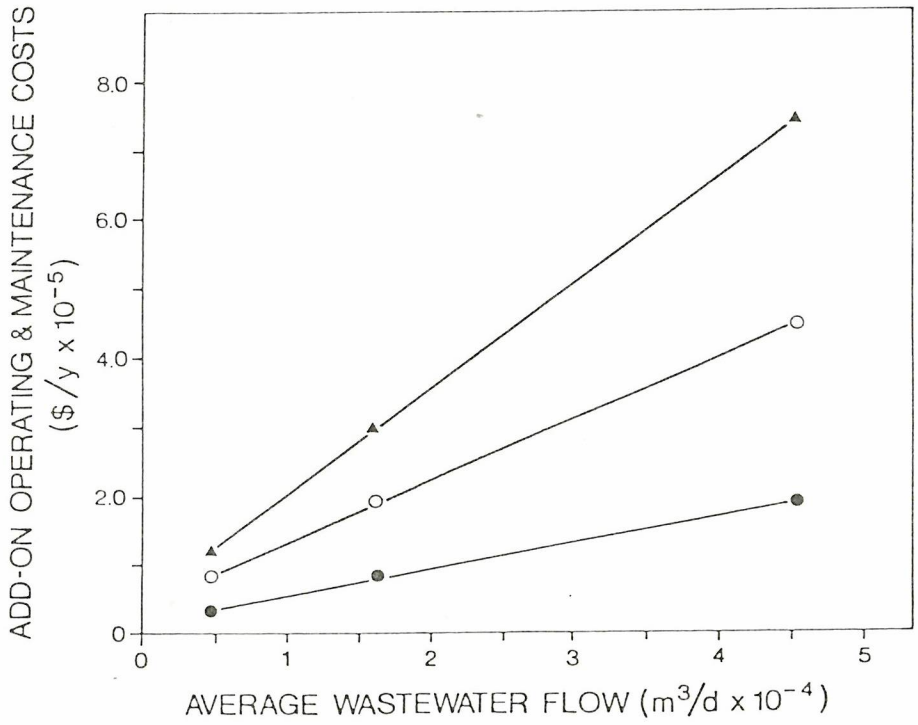


FIGURE 6. COSTS FOR "ADD-ON" CARBON AND NITROGEN REMOVAL PROCESSES

REFERENCES

Archer, J.D., "Summary Report on Phosphorus Removal", Canada-Ontario Agreement Research Report No. 83. Project No. 75-1-42, 1978.

Barnard, H.L. and Meiring, P.G.J., "Sources of Hydrogen Donors and Their Effects on Denitrifcation Rates", Progress in Water Technology 8, 577 (1977).

Christensen, M.H., Harremoes, P., "Biological Denitrification of Sewage: A Literature Review." Progress in Water Technology 8, 509 (1977).

Culp, G., Berjes, H.H., Puntenny, J., "Evaluating Wastewater Facility Staffing Needs." Jour. Water Pollution Control Federation 49, 2226 (1977).

CH_2M-Hill Ltd., "Summary Report Penticton Wastewater Disposal Plan", (1977).

CH_2M-Hill Inc., "Estimating Staffing for Municipal Wastewater Treatment Facilities." Report for US EPA, NTIS Pub. No. PB227-321/7 (1973).

Patterson, W.L. and Banker, R.F., "Estimating Costs and Manpower Requirements for Conventional Wastewater Treatment Facilities." Black and Veach for U.S. EPA NTIS PB 211-132 (1971).

Schmidke, N.W. and Cohen, D.B. "Municipal Sludge Disposal on Land, A Down to Earth Solution." Proc. 29th Annual Western Canada Water and Sewage Conference, Edmonton, 27 (1977).

Schmidt, J.W. "Advanced Wastewater Treatment Technology in Canada." Environmental Protection Service Report EPS-4-WP-79-5 (1979).

Sutton, P.M., Murphy, K.L., Jank, B.E. and Monaghan, B.A. "Single Sludge Nitrogen Removal Systems, Canada-Ontario Agreement Res. Report No. 88, Proj. No. 75-3-21 (1979).

U.S. EPA "Process Design Manual for Phosphorus Removal." (1976).

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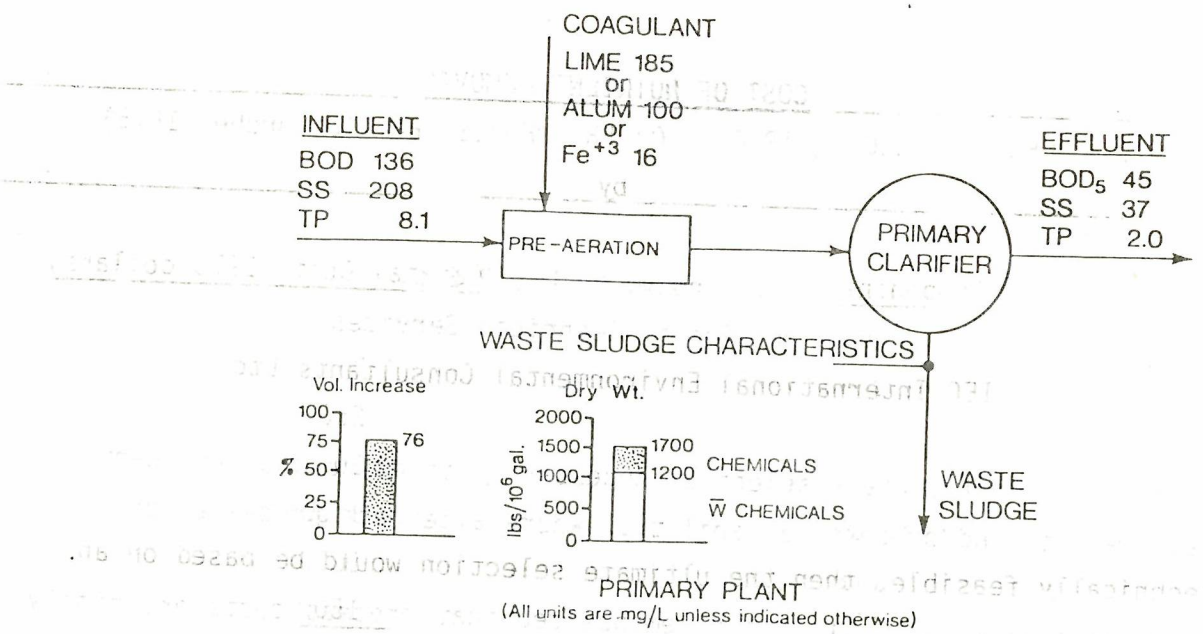
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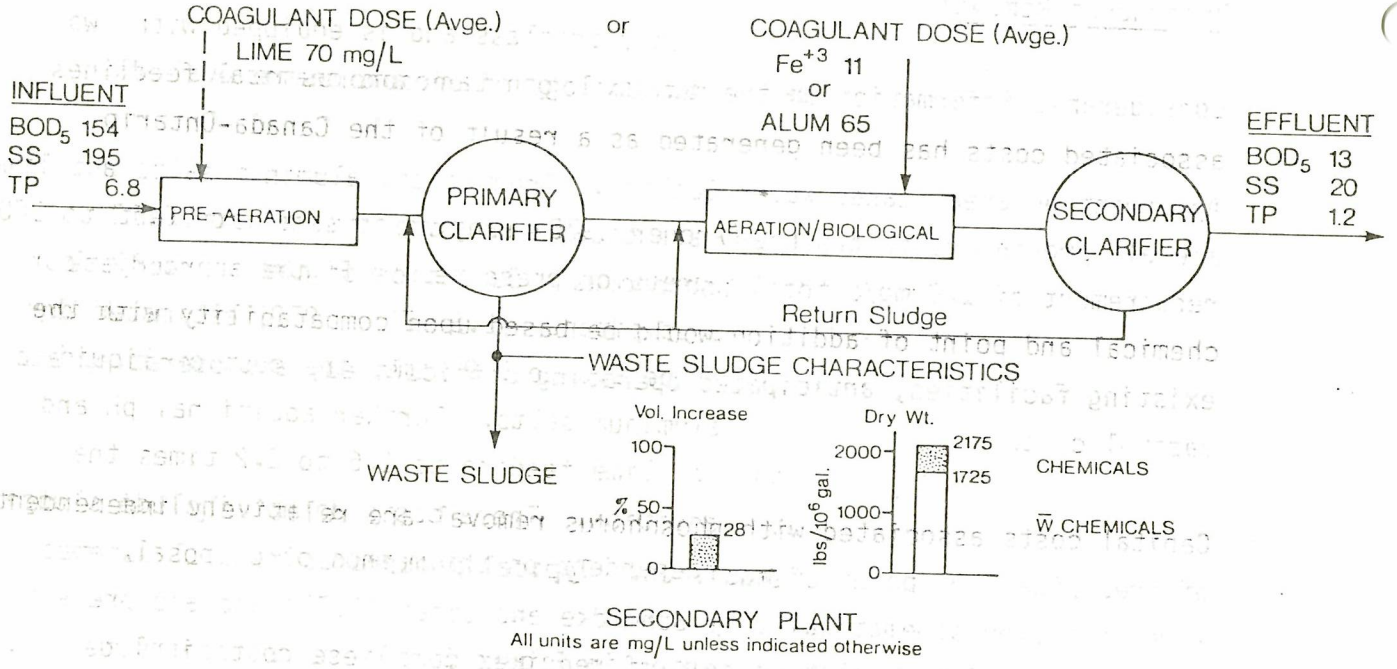


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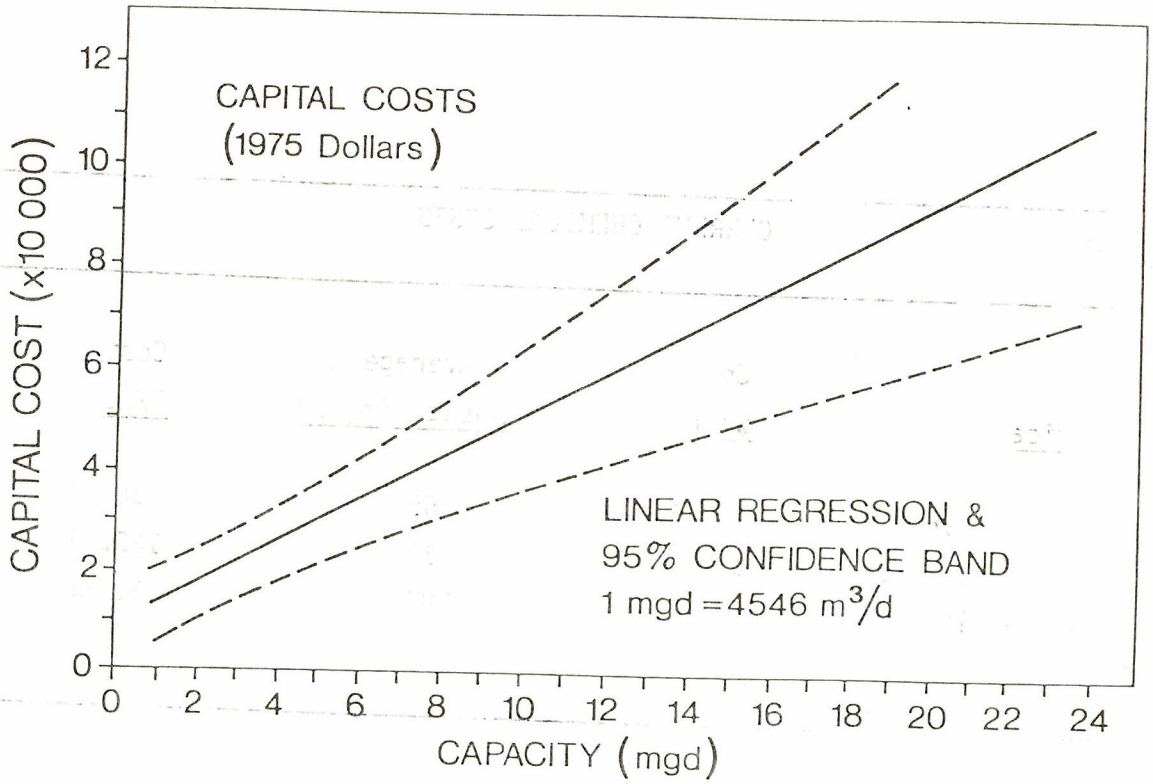
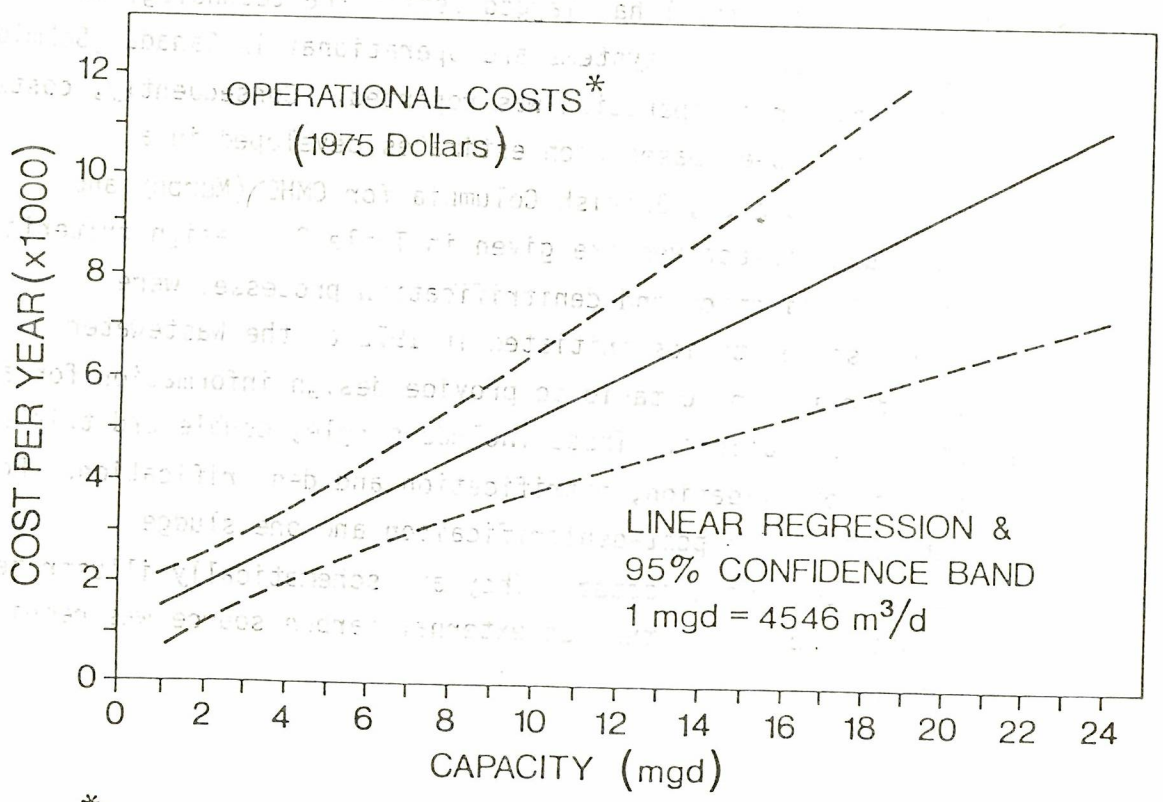


FIGURE 2. PHOSPHORUS REMOVAL EQUIPMENT-CAPITAL COSTS



*Costs Exclude Chemical Costs

FIGURE 3. PHOSPHORUS REMOVAL-OPERATIONAL COSTS

TABLE 2: CURRENT CHEMICAL COSTS

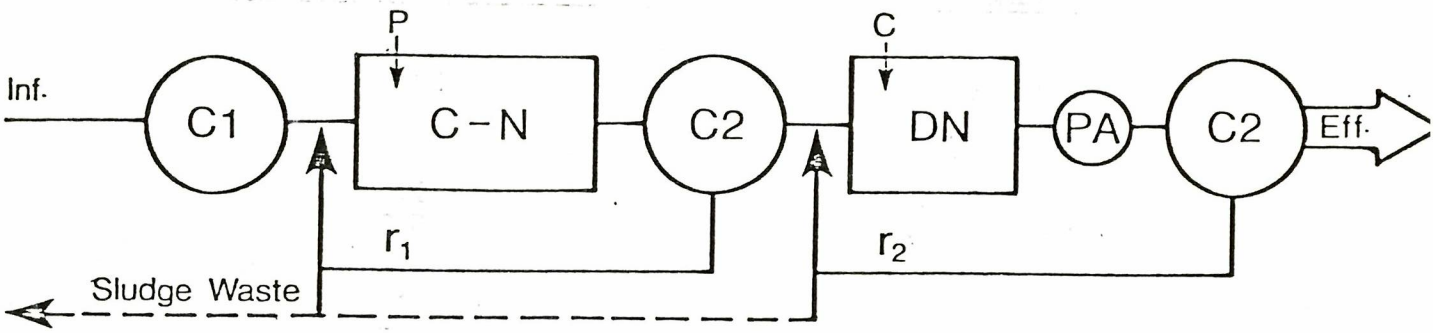
Chemical	Cost $/ton	Average Dosage (mg/L)	Cost $/MC
Alum (liquid)	84.50	65	56.05
Ferric Chloride	507.00	17	125.37
Hydrated lime	37.64	180	33.88

Nitrogen Removal

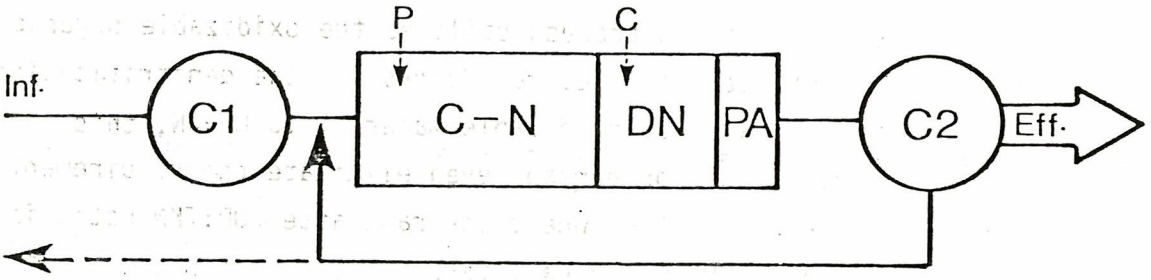
The technology of nitrogen removal has lagged behind the technology of phosphorus removal. No municipal systems are operational in Canada (Schmidt 1979), and only one industrial operation was reported. Consequently, costs for nitrogen removal have been based upon estimates developed in a feasibility study for Penticton, British Columbia for CMHC (Murphy and Wilson, 1978). Effluent objectives are given in Table 3. Design criteria for the biological nitrification and denitrification processes were developed from pilot scale studies initiated in 1972 at the Wastewater Technology Centre, Burlington, Ontario to provide design information for a variety of system configurations. These include single, double and triple sludge sytems for carbon oxidation, nitrification and denitrification. For Penticton, one and two sludge post-denitrificaiton and one sludge pre-denitrification system were costed. They are schematically illustrated in Figure 4. It should be noted that an external carbon source was required in all three systems.

Figure 4
Schematic of Nitrogen Removal Processes

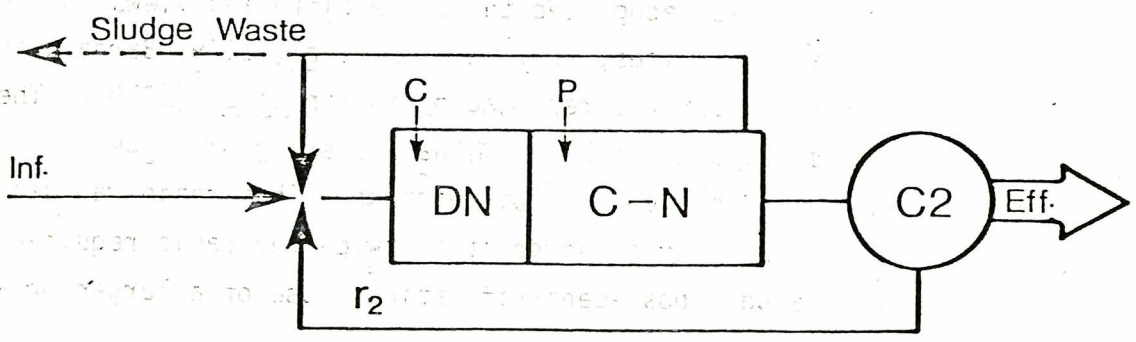
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Pickle Liquor	$2.11/100 kilograms
Equipment Replacement and Supply	0.5% total capital cost/year

TABLE 5: ECONOMIC COMPARISON OF BIOLOGICAL NITROGEN REMOVAL PROCESSES

	Activated Sludge	Nitrification	Two-Sludge Post-DN	One-Sludge Post-DN	One-Sludge Pre-DN
Present Worth[a]					
Capital	$ 2,618,000	$ 3,268,000	$ 4,802,000	$ 4,180,000	$ 4,217,000
Present worth of annual O&M	1,103,000	1,679,000	3,366,000	2,996,000	2,567,000
Total present worth	$ 3,721,000	$ 4,947,000	$ 8,168,000	$ 7,176,000	$ 6,784,000
Comparative rating	0.55	0.73	1.20	1.06	1.00
Annual Cost					
Amortized capital[a]	$ 318,000	$ 397,000	$ 583,000	$ 508,000	$ 512,000
O&M	134,000	204,000	409,000	364,000	312,000
Total annual cost	$ 452,000	$ 601,000	$ 992,000	$ 872,000	$ 824,000
Comparative rating	0.55	0.73	1.20	1.06	1.00

[a] Based on a 10.5-percent discount/interest rate for 20 years

NOTE: All costs are in January 1979 dollars

TABLE 6: COST COMPARISON OF BIOLOGICAL NITROGEN REMOVAL WITH OTHER METHODS

	Activated Sludge (x $1,000)	Nitrification (x $1,000)	Biological Nitrogen Removal (x $1,000)			Other Methods (x $1,000)		
			Two-Sludge Post-DN	One-Sludge Post-DN	One-Sludge Pre-DN	Clinoptiolite Ion Exchange	Ammonia Stripping	Land Disposal
Capital	$ 7,513	$ 8,600	$10,134	$ 9,512	$ 9,558	$15,980	$12,561	$19,434
Annual O&M	421	485	600	555	503	629	679	162
Present Worth of O&M[a]	3,456	3,992	4,938	4,568	4,140	5,177	5,589	1,333
Total Present Worth	$10,969	$12,592	$15,072	$14,080	$ 13,698	$21,157	$18,150	$19,712[b]
Comparative Rating	0.80	0.92	1.10	1.03	1.00	1.54	1.32	1.44

NOTE: Costs are in January 1979 dollars for a complete treatment facility at Penticton, B.C. with capacity of 1.8×10^4 m^3/d, 4540 kg BOD/d, and average annual flow of 1.55×10^4 m^3/d.

[a] Based on a discount rate of 10.5 percent for 20 years.

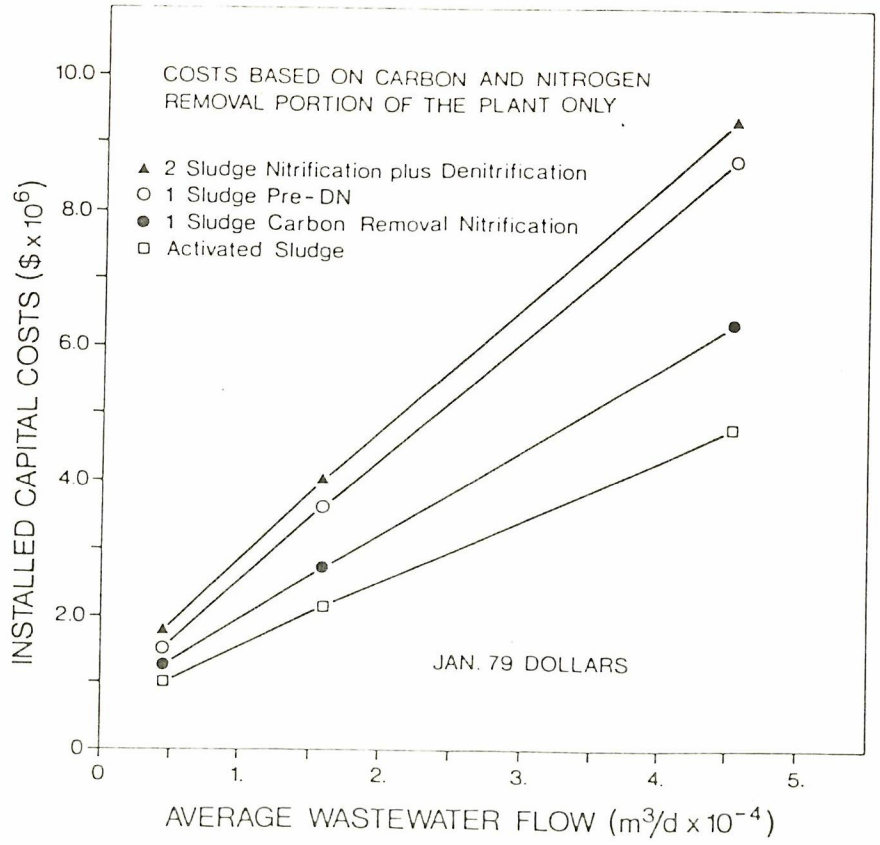
[b] Includes pipeline salvage value of $365,000 based on straight-line depreciation and a 40-year pipeline life, land salvage value of $690,000 based on 3-percent compound appreciation in land value for 20 years, and 10.5-percent discount rate.

alone and activated sludge treatment. The curves were prepared from separate cost estimates for a 0.45 x 10^4 m^3/d(1.0 IMGD), a 1.5 x 10^4 m^3/d (3.4 IMGD) and a 4.5 z 10^4 m^3/d (10 IMGD) plant using the wastewater characteristics at Penticton as a design basis. Equipment, power, chemical and labour costs are also based on the Penticton values.

Capital and O & M costs for new facilities presented in Figure 5 include only those portions of the treatment plants associated with carbon removal and nitrogen removal. This excludes headworks, primary clarifiers, sludge processing and disinfection. The capital costs in Figure 5 do not include the 20% contingency included in the costs in Table 5 but do allow for yardwork and engineering.

Figure 6 presents curves for the costs of "add-on" facilities to existing activated sludge treatment plants. The add-on capital costs were prepared by estimating the cost of additional basin volume, aeration, mixing and chemical feed equipment. It was assumed that no major renovations would be necessary to the existing final clarifiers and that they would be able to handle the increased solids loading caused by plant conversion. As before, the capital cost includes yardwork and engineering. A 15% contingency was added due to the uncertainties of predicting costs for renovations to existing plants.

Figure 5 shows that a 10 fold increase in plant size gives a 5 fold increase in capital cost for an activated sludge system and a 6 fold increase for a pre-denitrification system. The difference is caused by the relative difference in contribution to capital cost of the mechanical versus the structural components. For activated sludge systems, mechanical components represent a significantly higher percentage of total capital costs in comparison to pre-denitrification which requires more basin volume. Unit costs for mechanical components increase with plant capacity more slowly than structural components. Therefore, the higher percentage of mechanical components in an activated sludge plant is reflected in a lower rate of increase in total capital cost with increasing plant size.



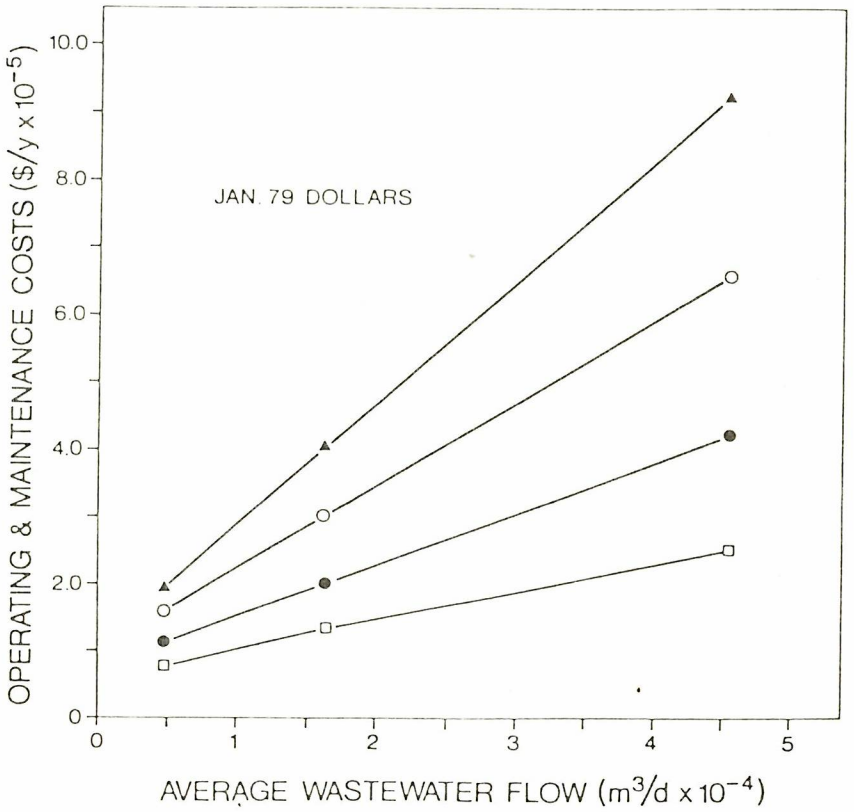
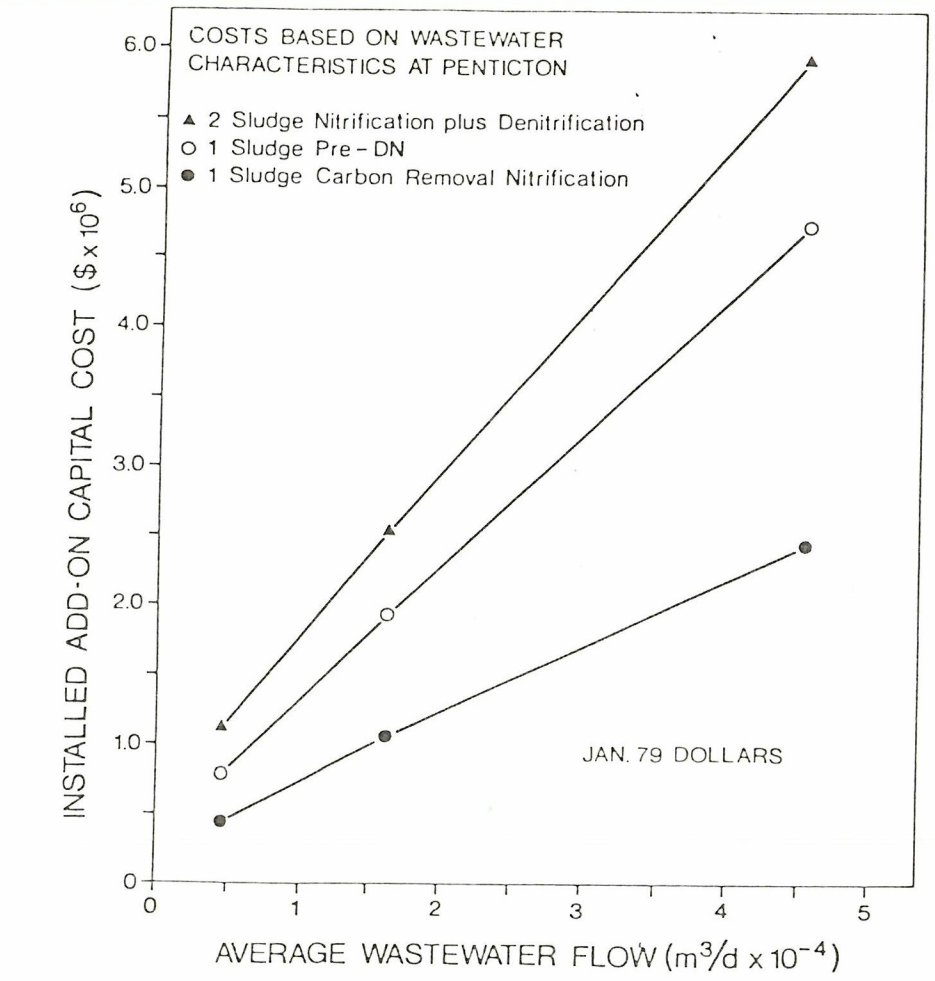


FIGURE 5. COSTS FOR CARBON AND NITROGEN REMOVAL PROCESSES



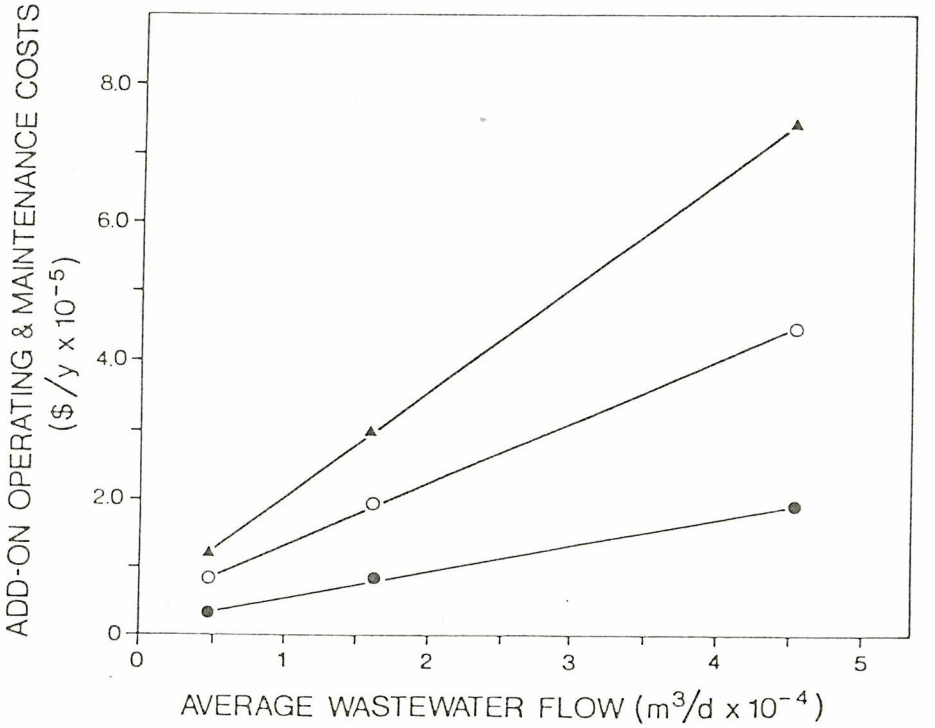


FIGURE 6. COSTS FOR "ADD-ON" CARBON AND NITROGEN REMOVAL PROCESSES

REFERENCES

Archer, J.D., "Summary Report on Phosphorus Removal", Canada-Ontario Agreement Research Report No. 83. Project No. 75-1-42, 1978.

Barnard, H.L. and Meiring, P.G.J., "Sources of Hydrogen Donors and Their Effects on Denitrifcation Rates", Progress in Water Technology 8, 577 (1977).

Christensen, M.H., Harremoes, P., "Biological Denitrification of Sewage: A Literature Review." Progress in Water Technology 8, 509 (1977).

Culp, G., Berjes, H.H., Puntenny, J., "Evaluating Wastewater Facility Staffing Needs." Jour. Water Pollution Control Federation 49, 2226 (1977).

CH_2M-Hill Ltd., "Summary Report Penticton Wastewater Disposal Plan", (1977).

CH_2M-Hill Inc., "Estimating Staffing for Municipal Wastewater Treatment Facilities." Report for US EPA, NTIS Pub. No. PB227-321/7 (1973).

Patterson, W.L. and Banker, R.F., "Estimating Costs and Manpower Requirements for Conventional Wastewater Treatment Facilities." Black and Veach for U.S. EPA NTIS PB 211-132 (1971).

Schmidke, N.W. and Cohen, D.B. "Municipal Sludge Disposal on Land, A Down to Earth Solution." Proc. 29th Annual Western Canada Water and Sewage Conference, Edmonton, 27 (1977).

Schmidt, J.W. "Advanced Wastewater Treatment Technology in Canada." Environmental Protection Service Report EPS-4-WP-79-5 (1979).

Sutton, P.M., Murphy, K.L., Jank, B.E. and Monaghan, B.A. "Single Sludge Nitrogen Removal Systems, Canada-Ontario Agreement Res. Report No. 88, Proj. No. 75-3-21 (1979).

U.S. EPA "Process Design Manual for Phosphorus Removal." (1976).