OPTIMIZATION OF BIOLOGICAL NUTRIENT REMOVAL IN A PILOT-SCALE TRICKLING FILTER-ACTIVATED SLUDGE PROCESS

by

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ABSTRACT

A pilot-scale study was undertaken to further the development of a combined trickling filter-activated sludge process designed for biological phosphorus removal and nitrification-denitrification. The system is called the FGR-SGR (fixed growth reactor-suspended growth reactor) process. The innovative aspect of the FGR-SGR process is the incorporation of a fixed growth (trickling filter) component into the conventional suspended growth (activated sludge) biological nutrient removal treatment train.

The objective of the study was to extend and optimize design and operational criteria for biological nutrient removal in the FGR-SGR process, including an assessment of optimum process reactor hydraulic retention times, internal recycle flow rates, and operating mixed liquor suspended solids (MLSS) concentration. Two pilot-scale processes were operated in parallel, to compare controlled changes in design parameters.

Both pilot plants consistently produced an effluent typically containing 10-15 mg/L suspended solids, less than 10 mg/L BOD_5, less than 0.01 mg N/L ammonia, and 2-3 mg N/L total kjeldahl nitrogen, regardless of design and operational changes. Nitrification in the FGRs accounted for greater than 85% of the process total nitrification, and the fixed growth nitrification was found to be first order with respect to ammonia concentration. A higher FGR irrigation (recycle) rate was associated with a significantly greater nitrification rate.

On the other hand, phosphorus removal was highly dependent on design and operational changes to the process. Daily short-term increases in the FGR recycle rate to prevent excess solids buildup on the media greatly improved phosphorus removal. With daily pulse hydraulic loading, low effluent orthophosphate concentrations (less

than 0.3 mg P/L) were observed when the ratio of the mass of volatile fatty acids (VFA) taken up in the anaerobic reactor to process influent total phosphorus (P) concentration was greater than 6 mg HAc/mg P; at lower ratios, effluent orthophosphate concentrations increased to greater than 1 mg P/L. The mass of volatile fatty acids taken up in the anaerobic reactor depended on the anaerobic actual hydraulic retention time, the steady-state process influent VFA concentration, and the steady-state mean MLSS concentration. Biological phosphorus removal was significantly better at an aeration basin mean MLSS concentration of approximately 3,000 mg/L, compared to one of 2,000 mg/L.

Denitrification in the anoxic reactor was accompanied by bacterial uptake of orthophosphate. Phosphorus uptake in the anoxic reactor accounted for approximately 45% of overall process total phosphorus removal. Bench-scale batch tests showed that following the completion of denitrification, secondary phosphorus release occurred for the remainder of the anoxic phase. Allowing significant concentrations of VFA to reach the anoxic reactor induced phosphorus release during the first few minutes of denitrification, reducing net anoxic phosphorus uptake.

Bench-scale batch tests designed to simulate the effects of manipulating internal recycle flow rates to dampen hydraulic shocks typically caused by the peak daily load in full-scale plants indicated that manipulation of the recycle flows has the potential to improve phosphorus removal in the process.

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1. INTRODUCTION

The discharge of nutrient-rich wastes such as domestic sewage into environmentally-sensitive surface waters can cause a serious deterioration in water quality. Nutrient enrichment of surface waters (eutrophication) can cause algae blooms and promote the growth of nuisance aquatic vegetation. The nutrients of primary concern are phosphorus in the case of discharges to fresh water, and nitrogen in the case of marine discharges. In addition to eutrophication problems, ammonia nitrogen and nitrite nitrogen are toxic to fish.

Chemical removal of phosphorus is currently practiced, but chemical additions increase treatment plant operating costs and generate larger solids disposal volumes than biological systems. In many situations, biological phosphorus removal is therefore preferred. Current technology for biological phosphorus removal is generally based on activated sludge systems only; existing trickling filters are usually decommissioned and demolished or converted to other uses when activated sludge biological phosphorus removal systems are installed. Use of trickling filters as a key component in biological nutrient removal processes could, therefore, be beneficial.

Current technology for biological conversion of ammonia to nitrate (nitrification) is based on the oxidation of ammonia by either suspended growth bacteria (in activated sludge systems) or by fixed growth bacteria, which are cultured in a slime layer attached to some solid medium (e.g., in trickling filters and rotating biological contactors). Nitrification in activated sludge systems generally requires relatively long hydraulic retention times for the process liquid in the aeration basin, and the aeration basin is therefore relatively large. Nitrification in activated sludge systems also requires relatively long solids retention times (sludge ages), which can create a

poor-settling biomass, which in turn requires a larger secondary clarifier.

In activated sludge biological nutrient removal systems, both nitrification and bacterial phosphorus uptake-storage occur in a single aerated basin. However, the aerated retention time required for phosphorus uptake may differ from that required for nitrification. It may therefore not be possible to optimize the size of the aeration basin for both of the two biochemical processes simultaneously.

The study described in this thesis was designed to further develop a combined trickling filter-activated sludge system for biological phosphorus and nitrogen removal. The system is called the FGR-SGR (fixed growth reactor-suspended growth reactor) process. The study was focused on the design and operational characteristics of the FGR-SGR process.

In the FGR-SGR process, partial or complete nitrification can be accomplished by the fixed growth biomass attached to the FGR (trickling filter) media. Since nitrification is typically the factor which determines the aeration basin size in activated sludge biological nutrient removal plants, inclusion of the trickling filters should allow a reduction in the required size of the aeration basin (i.e., the aeration basin can be sized to achieve optimum bacterial phosphorus uptake). Other advantages of incorporating trickling filters into biological phosphorus removal technology include the option of retro-fitting existing trickling filter plants for biological nutrient removal, process stability added by the fixed growth bacteria in the trickling filters, low land use, and good sludge settling qualities.

The Objectives of this study are given in Chapter 2. The Literature Review (Chapter 3) includes a summary of current technology for coupled trickling filter-activated sludge sewage treatment systems, nitrification in trickling filter systems, denitrification in activated sludge systems, and biological phosphorus removal in activated sludge

systems. Materials and Methods are described in Chapter 4. The Results and Discussion are given in Chapters 5 and 6, respectively. Conclusions are contained in Chapter 7, and Recommendations for Further Research are outlined in Chapter 8.

2. OBJECTIVES

The purpose of this project was to extend and optimize design and operational criteria for enhanced biological phosphorus and nitrogen removal in the FGR-SGR process. The specific research objectives were as follows:

- investigation of the optimum hydraulic retention times for biological phosphorus removal in the suspended growth components of the FGR-SGR system, including an assessment of the possibility of reducing suspended growth aeration requirements by utilizing oxidized forms of nitrogen (rather than dissolved oxygen) for biological phosphorus uptake;

- investigation of the effects on biological phosphorus removal and nitrification of varying the internal recycle flow rates within the FGR-SGR process; and

- investigation of the optimum operating mixed liquor suspended solids concentration in the suspended growth component of the FGR-SGR system for biological phosphorus removal and nitrification-denitrification.

3. LITERATURE REVIEW

3.1 COMBINED TRICKLING FILTER-ACTIVATED SLUDGE PROCESSES

The history and development of combined trickling filter-activated sludge processes for the removal of oxygen demand and suspended solids, together with process modeling and design procedures, were reviewed previously (Gibb, 1990). In general, advantages of combining or coupling suspended and attached growth systems include protection against biomass washout, improved handling of industrial discharges or toxic shock loads, improved sludge settling properties, and overall ease of operation (USEPA, 1993). In a comprehensive review of full-scale combination trickling filter-activated sludge plants of all types in the U.S.A., Harrison et al. (1984) concluded that the combined systems generally offer an energy-efficient alternative to conventional activated sludge options, and that the combined processes are capable of producing a very high quality effluent.

The development of combined trickling filter-activated sludge processes in North America for the removal of oxygen demand, suspended solids, and in some cases nitrification, has resulted in three process schematics, the Trickling Filter/Solids Contact (TF/SC), Trickling Filter/Activated Sludge (TF/AS), and the Activated Biofilter/Activated Sludge (ABF/AS) processes. More recently, a combined trickling filter-activated sludge process designed to include biological phosphorus removal, called the Fixed Growth Reactor-Suspended Growth Reactor (FGR-SGR) process, was developed in British Columbia.

The TF/SC process, which incorporates trickling filters in series with a small aerated solids contact basin, has been developed and successfully applied at more than fifty

locations in North America; the TF/SC process is reported to be lower in capital, operating, and maintenance costs than competing activated sludge processes (Parker et al., 1990). In the TF/SC process, settled sewage is irrigated over the trickling filter media. The effluent from the trickling filter flows to a fully mixed, aerated activated sludge (solids contact) basin, and then to the secondary clarifier; settled biological solids from the secondary clarifier are returned to the solids contact basin (Matasci et al., 1986). The majority of soluble biochemical oxygen demand (BOD) removal occurs in the trickling filter, and the primary function of the solids contact basin is to increase solids capture and particulate BOD removal. The solids contact basin, which has a hydraulic retention time of one hour or less and a solids retention time of two days or less, is not designed to promote the growth of suspended nitrifiers (Parker et al., 1992). However, nitrification may occur in the solids contact basin if the trickling filter is designed for nitrification, since nitrifiers sloughed from the trickling filter will be active in the aerated solids contact basin. Therefore, nitrification in the solids contact basin can be sustained in the TF/SC process by fixed growth nitrifiers continuously sloughing from the trickling filter media (USEPA, 1993).

Similarly, in the TF/AS process, settled sewage is irrigated over the trickling filter media, the effluent from the trickling filter flows to an aerated activated sludge basin, and settled biological solids are returned to the aeration basin. The most common application of TF/AS in systems requiring nitrification is to design the trickling filter as a "roughing filter" for partial removal of BOD, and to design the activated sludge system with sufficient solid and liquid retention times so that nitrification can be sustained by suspended growth organisms. As in the TF/SC process, nitrification in the aeration basin can also be sustained by nitrifiers sloughing from the trickling filter media (USEPA, 1993).

The ABF/AS process has been installed in at least 43 locations in the U.S.A. (Arora

and Umphres, 1987). The ABF/AS system is similar to the TF/SC and TF/AS processes in that the effluent from the trickling filter flows to an aerated activated sludge basin, and then to the secondary clarifier. However, in contrast to the TF/SC and TF/AS processes, in the ABF/AS system, settled biological solids from the secondary clarifier are returned to mix with the primary clarifier overflow, and the mixture is irrigated over the trickling filter media (USEPA, 1993). The purpose of recycling biological solids over the trickling filter media is to increase the removal of oxygen demand by addition of the biological solids which slough from the media, allowing a higher organic loading than for systems without solids recycle (Arora and Umphres, 1987).

The FGR-SGR process was pioneered at a full-scale domestic wastewater treatment facility at Salmon Arm, British Columbia, in 1986. The design for the FGR-SGR process was developed from the principles for activated sludge-type (suspended growth) biological nutrient removal processes, and those developed for BOD removal and nitrification in combined trickling filter-activated sludge systems (Kelly, 1987). A study carried out during 1988-89 at the full-scale plant at Salmon Arm demonstrated that the FGR-SGR process could sustain the organisms required for enhanced biological phosphorus removal. Bacterial phosphorus uptake rates in the full-scale plant were competitive with those published for activated sludge-type systems, the average removal of phosphorus from the process influent over an eight month operating period was 7 mg P/L, and the process suspended solids consistently contained 4-5% phosphorus by dry weight, indicating a significant degree of bacterial phosphorus storage (Gibb et al., 1989; and Gibb, 1990). However, the difficulties associated with full-scale research and the need for the full-scale plant to meet permit requirements prevented the experimentation needed to develop design and operating criteria required for the next generation of full-scale plants. Further, the Salmon Arm study did not include nitrogen removal. It was subsequently concluded that pilot-

scale investigations were the best method for developing the required data.

A pilot-scale study was designed in 1990 to refine and further develop design criteria for biological phosphorus removal and nitrification-denitrification in the FGR-SGR process (Gibb et al., 1993). The results of that study form the basis of this thesis.

3.2 NITRIFICATION IN TRICKLING FILTER SYSTEMS

Nitrification is the biochemical oxidation of ammonia nitrogen (NH_4^+) to nitrate nitrogen (NO_3^-) by autotrophic bacteria under aerobic conditions. The two species of bacteria of most importance in nitrification are *Nitrosomonas*, which convert ammonia to nitrite (NO_2^-), and *Nitrobacter*, which convert nitrite to nitrate. Other species of nitrifying bacteria include *Nitrosospira*, *Nitrosolobus*, and *Nitrosovibrio*. Nitrifiers are slow-growing organisms, and are therefore particularly susceptible to toxicants; however, nitrifiers are capable of adapting to toxic substances which are consistently present (Halling-Sorensen and Jorgensen, 1993).

In trickling filters, nitrifiers grow in a slime layer or film attached to a solid media. Many types of media have been used in trickling filters, including rocks, horizontal wood slats, random plastic rings, polyethylene strips, and corrugated plastic sheets. Corrugated plastic sheets, which are commonly used in modern trickling filters, fall into two classifications, vertical and cross-flow; of the two, cross-flow media has a higher oxygen transfer efficiency and a higher contact time between the biofilm and the process liquid, and is therefore preferred for nitrification systems (Parker and Merrill, 1984).

Empirical and mechanistic models for substrate removal in trickling filters were reviewed previously (Gibb, 1990). Complications in modeling trickling filter performance include the influence of particulate organics, and non-homogeneous

biofilm thickness and density. Mechanistic biofilm models, which usually include both microbial kinetics and substrate diffusion through the biofilm, consist of a set of partial differential equations, the solution of which can be infeasible even with the best numeric techniques. Accordingly, design of attached growth nitrification processes is primarily based on empirical data from pilot-scale and full-scale systems. However, empirical data are generally consistent with mechanistic biofilm models (USEPA, 1993).

According to Halling-Sorensen and Jorgensen (1993), attached growth systems generally oxidize more nitrogen per unit of bed volume than activated sludge processes. The degree of nitrification achieved in trickling filters depends on a number of factors, including the BOD loading rate, temperature, dissolved oxygen concentration, ammonia concentration, and the pH and alkalinity of the process liquid. Many of the factors are interrelated. At high BOD concentrations, factors which affect nitrification in trickling filters include hydraulic loading, hydraulic pattern and retention time on the filter media, the dissolved oxygen concentration in the process liquid, pH, temperature, feed total kjeldahl nitrogen (TKN) concentration, and feed BOD concentration (USEPA, 1993). According to Gullicks and Cleasby (1986), empirical design curves for nitrification in trickling filters should incorporate parameters for hydraulic loading, influent ammonia concentration, the effects of recycling, and wastewater temperature.

High BOD loading rates tend to suppress nitrification, since autotrophic nitrifying bacteria are out-competed by faster growing heterotrophs. The presence of organic matter inhibits the growth of nitrifiers, due to increased competition for dissolved oxygen by heterotrophs (Halling-Sorensen and Jorgensen, 1993). The BOD load on the trickling filter therefore affects the ratio of heterotrophs to autotrophs in the biofilm (Andersson et al., 1993).

Reduced temperatures are known to result in slower intrinsic bacterial nitrification rates. According to USEPA (1993), the relative effect of temperature on nitrification in trickling filters depends on BOD loading, oxygen availability, influent and effluent ammonia concentration, and hydraulic loading conditions. However, substrate removal in biofilms is frequently limited by substrate mass transfer (either oxygen or ammonia) rather than by the intrinsic biochemical reaction rate, since the surface area to volume ratio of the biomass is 10-100 times smaller than for suspended biomass (Siegrist and Gujer, 1987). Therefore, biofilms are generally less sensitive to temperature changes than suspended growth cultures, where the substrate removal rate is limited by the intrinsic biochemical reaction rate (Halling-Sorensen and Jorgensen 1993). The Arrhenius relationship is often proposed to describe temperature effects on fixed film biokinetics; however, if the substrate removal rate is limited by substrate mass transfer, this relationship is not appropriate. In a review of the data from eight nitrifying trickling filter facilities, Okey and Albertson (1989a) reported that temperature effects on the intrinsic biochemical reaction rate for nitrification in trickling filters are masked by oxygen diffusion limitations in the ammonia concentration ranges in which ammonia flux is not limiting.

For the most part, trickling filters for nitrification are used as a separate-stage tertiary treatment step, to treat the clarified effluent from a secondary treatment process. According to Boller and Gujer (1986), since tertiary nitrification produces only 2-3 mg of total suspended solids (TSS)/L, no additional clarifier is required. Single-stage trickling filter systems, where carbon oxidation and nitrification are accomplished in a single unit, are relatively rare. Two-stage systems generally yield a higher quality effluent than single stage systems; however, single-stage systems can be more cost-effective, requiring fewer unit operations. Further study is required to directly compare single-stage and tertiary trickling filter systems for nitrification (USEPA, 1991). The FGR-SGR process most closely resembles a single-stage system, since the

FGRs are located upstream of the secondary clarifier.

A recent study of nitrification in single-stage trickling filter systems identified only ten such plants operating in the U.S.A., with six of these using the TF/SC process. It was found that, to allow nitrification to effluent ammonia levels of less than 4 mg N/L, organic loading in single-stage plants should be less than 3.2 g $BOD_5/m^2/d$, and that effluent ammonia levels of 2-4 mg N/L will generally be reached when effluent BOD_5 concentration is less than 15 mg/L, and preferably less than 10 mg/L (USEPA, 1991).

According to Parker and Richards (1986), nitrification in single-stage trickling filter systems begins only when the bulk solution soluble BOD_5 concentration is 20 mg/L or less; therefore, in single-stage systems, nitrifiers can only become established in the lower portions of the tower, where BOD concentrations are low. On the other hand, Gullicks and Cleasby (1990) reported that good nitrification performance was observed in a pilot-scale trickling filter at sustained soluble COD concentrations of 60-66 mg/L; they concluded that the bulk liquid dissolved oxygen concentration had a much greater impact on nitrification that soluble COD concentration; and that carbonaceous oxygen demand had a greater impact on nitrification at low hydraulic loads than at high hydraulic loads.

In second-stage (tertiary) trickling filters, increasing the ammonia loading rate (by either increasing the ammonia concentration or the hydraulic loading rate) generally increases the nitrification rate; however, the percent removal efficiency is reduced (Gullicks and Cleasby, 1986). According to Boller and Gujer (1986), the nitrification rate in biofilms depends on ammonia concentration in the low concentration ranges (less than 3-5 mg N/L), and is zero order with respect to ammonia concentration at higher ammonia concentrations, where the nitrification rate is limited by oxygen diffusion. A review of several pilot-scale studies and one full-scale study of tertiary

trickling filter nitrification by Okey and Albertson (1989a) supported the findings of Boller and Gujer; according to Okey and Albertson, the regime that is zero order for ammonia is half order for oxygen; at low ammonia concentrations (less than 2-5 mg N/L), the rate is first order with respect to ammonia. Similarly, Parker and Richards (1986) reported that nitrification resembles a zero-order reaction with respect to ammonia concentration, at least down to concentrations of 2 mg N/L.

Diurnal variations in ammonia loading can affect trickling filter performance in systems designed for complete nitrification; if ammonia reaches the lower depths of the media only during peak loading hours, biofilm development in the lower depths may be patchy. This difficulty can be overcome by operating two trickling filters in series; every 7-14 days the order of the two reactors is reversed, ensuring uniform biofilm development in both reactors (Gujer and Boller, 1984 and Andersson et al., 1993).

It has been well-documented that nitrification rates decline significantly if the pH drops below the neutral range. In general, the optimum range of pH for nitrification is regarded to be 6.5-8.0 (USEPA, 1993). The nitrification reaction results in approximately 7.1 mg alkalinity (as $CaCO_3$) destroyed per mg of ammonia oxidized (USEPA, 1993). Therefore, in low-alkalinity waters, as nitrification proceeds, the pH may drop and inhibit the nitrifying bacteria. According to Gujer and Boller (1984), nitrification in trickling filters becomes alkalinity-limited if residual alkalinity falls below 1 meq/L, due to accumulation of the products of respiration (which includes hydrogen ions) within the biofilm. Huang et al. (1989) reported that nitrification in a laboratory-scale tertiary trickling filter system was severely reduced when the influent alkalinity to ammonia nitrogen concentration was less than 7:1.

Recirculation of the trickling filter effluent over the media may or may not be practiced. A higher rate of recycle results in a higher hydraulic loading to the media.

According to USEPA (1991), recirculation is beneficial to trickling filter performance, lowering the applied substrate concentrations, assuring uniform surface wetting, and helping to control filter flies and other biofilm predators. However, in a review of eight tertiary trickling filter facilities, Okey and Albertson (1989a) found little or no effect on nitrification derived from varying the hydraulic loading rate between 0.34-1.70 L/m²/s, provided that the ammonia loading was kept below 1.2 g N/m²/d. At loadings greater than 1.2 g N/m²/d, limited data indicated that low hydraulic loading rates resulted in suppressed nitrification rates (Okey and Albertson, 1989b)

Mass invasions of higher organisms which graze on the nitrifying biomass in trickling filters (e.g., filter fly larvae, worms, and snails) can result in reduced nitrification rates. In a comprehensive pilot-scale investigation of trickling filter nitrification using high-density media (230 m²/m³), Boller and Gujer (1986) reported that a hydraulic loading rate of at least 3 m/hr (0.83 L/s/m²) was necessary to guarantee complete wetting of the media, which in turn controlled predation by higher organisms. Parker et al. (1989) found that weekly flooding of the media was an effective means of controlling filter fly larvae, but not worms or snails.

Since nitrifiers are slow-growing organisms, a lengthy period may be required after start-up of a plant to achieve steady-state operation. In a pilot-scale study of tertiary (second-stage) trickling filter nitrification in Switzerland, Gujer and Boller (1984) reported that more than 400 days were required to reach steady-state performance, and that the previous loading record of a trickling filter influenced its performance over a long period of time.

Recent evidence indicates that simultaneous nitrification and denitrification occurs in biofilm reactors. According to Watanabe et al. (1992), nitrifiers and denitrifiers co-exist throughout the biofilm in fixed growth reactors, and the activities of the two types of organisms depend on the dissolved oxygen level within the biofilm.

3.3 DENITRIFICATION IN ACTIVATED SLUDGE PROCESSES

Biological denitrification is the reduction of nitrate and nitrite to gaseous forms of nitrogen, by a relatively broad range of facultative heterotrophic bacteria which can use nitrogen-bound oxygen in the absence of free dissolved oxygen as an electron acceptor. Denitrification is commonly thought to occur only in the presence of nitrate and the absence of molecular oxygen (i.e., under anoxic conditions), since the presence of oxygen tends to repress the synthesis and inhibit the activity of the bacterial enzymes required for denitrification. Lie and Welander (1994) developed a linear relationship between dissolved oxygen and oxidation reduction potential in activated sludge denitrification systems; based on an extrapolation of the relationship, they concluded that dissolved oxygen had a negative effect on denitrification rates, even at concentrations of less than 0.1 mg/L. They also found that the denitrification rate decreased linearly with increasing oxidation reduction potential, but that the effect differed among sludges from different treatment plants.

Enzyme synthesis for denitrification takes 0.5 hours or more, while enzyme activity can be stopped or reduced and then re-activated within minutes, depending on the presence or absence of the inhibitor (von Schulthess et al., 1994). There is some evidence that denitrification occurs in aerobic systems (USEPA 1993, and Halling-Sorensen and Jorgensen 1993). However, it may be difficult to distinguish between true aerobic denitrification and denitrification in anoxic microniches within the biomass (von Schulthess et al., 1994).

Denitrification requires a carbon source; carbon may be provided by the endogenous decay of the process biomass, by adding a supplemental carbon source (usually methanol), or by the organic material present in the raw wastewater. The concentration and nature of the available organic carbonaceous substrates has been shown to affect the denitrification rate in biological treatment systems (e.g., Carley

and Mavinic, 1991). In a bench-scale sequencing batch reactor system, Tam et al. (1994) found that acetate and propionate were effective in enhancing denitrification rates in activated sludge. Lynga and Balmer (1992) found that the denitrification rate was controlled by nitrate concentration at COD:TKN ratios of greater than 15:1, and by the supply of easily degradable carbon at COD:TKN ratios of less than 15:1. Since the process in this study was designed to use the raw wastewater BOD as a carbon source, endogenous activity and supplemental carbon additions were not considered here.

In single-sludge wastewater treatment systems, stoichiometric relationships show that, of the 4.6 g of oxygen required to oxidize 1 g of ammonia nitrogen to nitrate nitrogen, 2.86 g of oxygen can be recovered during the reduction of 1 g of nitrate nitrogen to nitrogen gas. The oxygen requirement can therefore be theoretically reduced by up to 63%, if denitrification using the carbon in the raw wastewater is undertaken (USEPA, 1993).

Denitrifiers are inhibited by pH outside the range 6-8. Since the denitrification reaction theoretically produces 3.57 mg alkalinity (as $CaCO_3$) per mg nitrate nitrogen reduced, approximately half of the alkalinity lost during nitrification is replaced, helping to buffer system pH changes (USEPA 1993, and Halling-Sorensen and Jorgensen 1993).

Denitrification occurs in the temperature range 5-35°C. An Arrhenius-type function is often used to predict the effect of temperature on denitrification rates (USEPA, 1993). However, since many denitrifying organisms are adaptive to temperature changes, there may be a difference between short-term and long-term temperature effects (Halling-Sorensen and Jorgensen, 1993).

The sludge blanket in the secondary clarifier can work as a denitrifying reactor,

substantially increasing nitrogen removal (Siegrist and Gujer, 1994). Further, solids residence times of at least 0.5 hr in the secondary clarifier sludge blanket provide an opportunity for bacterial synthesis of the enzymes required for nitrate reduction, improving denitrification in the anoxic reactor (Wild et al., 1994).

3.4 BIOLOGICAL PHOSPHORUS REMOVAL IN ACTIVATED SLUDGE PROCESSES

Enhanced biological phosphorus removal in activated sludge waste water treatment processes, including modeling, design, operation, control, and performance, were reviewed previously (Gibb, 1990). The general fundamental requirements for biological phosphorus removal to occur are outlined below.

- An anaerobic (no dissolved oxygen present) - aerobic (dissolved oxygen present) sequence is required in the suspended growth bioreactor. Since electron acceptors (molecular oxygen and nitrate-bound oxygen) are absent in the anaerobic phase, metabolism of carbon substrates via the electron transport chain is not possible. However, according to current biochemical models (e.g., reviewed by Wentzel et al., 1991a and 1991b), phosphorus-storing bacteria in the anaerobic phase are able to take up short-chain volatile fatty acids (e.g., acetic acid) from the anaerobic bulk solution, and convert fatty acids to poly-ß-hydroxyalkanoates (PHA), a high-energy storage product. Biochemical conversion of acetate to PHA is an energy-demanding reaction; the required energy is thought to be provided by the hydrolysis of adenosine triphosphate (ATP) to adenosine diphosphate (ADP), resulting in the liberation of orthophosphate ions (PO_4^{3-}). Regeneration of ATP from ADP is thought be accomplished by transfer of an energy-rich phosphoryl group from a polyphosphate (polyP) storage product within the cell. Thus short-chain volatile fatty acids (VFA) are taken up and converted to PHA under anaerobic conditions, with the simultaneous degradation of intracellular polyP.

Orthophosphate ions liberated during the hydrolysis of ATP are transported across the cell membrane to the bulk solution. According to Wentzel et al. (1991b), the theoretical molar ratio of phosphorus release to acetate uptake to is 1:1; according to Comeau et al. (1987), the ratio is approximately 1.5:1. Under subsequent aerobic conditions, the bacteria oxidize their PHA reserves, using part of the resulting energy to rebuild intracellular polyP reserves. Uptake and storage of phosphorus under aerobic conditions can exceed anaerobic phosphorus release, to the point where all of the released phosphorus is taken up, plus some or all of the phosphorus present in the process influent. Many researchers have shown that the amount of phosphorus taken up in the aerobic phase is proportional to the amount of phosphorus released in the anaerobic phase (e.g., Abu-ghararah and Randall, 1991 and Okada et al., 1991).

- Following the aerobic phase, the liquid and solid phases must be separated (usually be gravity settling), and the phosphorus-rich biosolids returned to the anaerobic reactor to begin another cycle.

- An adequate supply of short-chain organic acids must be supplied to the anaerobic zone of the bioreactor. The concentration and nature of soluble carbon-based compounds provided to the anaerobic bulk solution has been shown to affect the degree of biological phosphorus removal; increasing the concentration of the products of fermentation (mainly acetic acid, propionic acid, and butyric acid - often grouped as volatile fatty acids) in the anaerobic zone generally increases the degree of biological phosphorus removal (Malnou et al., 1984; Oldham, 1985; and Lotter and Pitman, 1992).

- Processes which include nitrification must also include denitrification, since the presence of nitrates as an electron acceptor in the anaerobic phase is thought to allow denitrifying bacteria to oxidize volatile fatty acids, reducing the amount of

easily-degradable carbon available, and suppressing the carbon storage-phosphorus release part of the biological phosphorus removal mechanism.

- Phosphorus must be removed from the system by wasting a portion of the phosphorus-rich biosolids (usually from the aerobic reactor or return settled biosolids flow).

- The solids retention time in the secondary clarifier sludge blanket should not be excessive. Anaerobic conditions in the sludge blanket promote endogenous phosphorus release by the process bacteria, increasing final effluent phosphorus concentrations (Oldham, 1985).

Many researchers have observed that phosphorus release continues at a slower rate following the disappearance of VFA from the anaerobic bulk solution. There is evidence to show that, in addition to the VFA entering the anaerobic phase via the process influent, VFA are generated in the anaerobic reactor by fermentation of more complex substrates (Brodisch, 1985 and Meganck et al., 1985). Since the generation rate of VFA by fermenting bacteria is much slower than the rate of VFA uptake by phosphorus accumulating bacteria, the fermentation rate limits the rate of VFA uptake-phosphorus release, if no other source of VFA (e.g., VFA added to the anaerobic phase from some external source) is available (Wentzel et al., 1991a). In support of the above, Comeau et al. (1987) documented a shift from a fast to a slow rate of phosphorus release and bacterial PHA storage, as soon as VFA added to the anaerobic phase disappeared from solution. On the other hand, two rates of anaerobic VFA (acetic acid) uptake-phosphorus release in the presence of significant bulk-solution concentrations of acetic acid have been documented by some researchers, an initial, rapid reaction rate being followed by a slower rate (e.g., Wentzel et al., 1987 and Gibb, 1990). Wentzel et al. (1987) suggested that the slowing of the reaction rate occurred as bacterial polyP reserves declined below a certain value.

The species of bacteria responsible for biological phosphorus removal has yet to be established. Organisms of the genus *Acinetobacter* have often been found in significant concentrations in biological phosphorus removal processes, and have been shown to be capable of accumulating excess phosphate (e.g., Lotter, 1985 and Wentzel et al., 1991a). However, Wentzel et al. (1991a) noted that the procedure commonly used to enumerate organisms in activated sludge systems has recently been found to overestimate the *Acinetobacter* count. Other researchers have concluded that bacterial groups other than *Acinetobacter* are also capable of enhanced phosphorus removal (e.g., Brodisch and Joiner, 1983 and Suresh et al., 1985).

In a study of *Acinetobacter* isolated from the activated sludge in a BNR plant, Cloete and Bosch (1994) concluded that phosphate was mainly accumulated by smaller, slow-growing cells, and that the majority of phosphorus was accumulated during the lag growth phase. In a study of 156 *Acinetobacter* isolated from a biological nutrient removal pilot plant, Beacham et al. (1992) found that not all of the isolates were capable of accumulating phosphate, and that there was no relationship between the individual genospecies of *Acinetobacter* and the ability to accumulate phosphate; they concluded that attempts to model process performance by following population trends in particular genospecies are probably unfounded. However, Ubukata and Takii (1994) found that the bacteria isolated from a biological nutrient removal plant did not accumulate excess phosphorus unless subjected to at least two anaerobic-aerobic incubation cycles; therefore, the results of studies where isolates are cultured under strictly aerobic conditions might be misleading.

According to the results of enhanced culture studies by Wentzel et al. (1991a), the bacteria responsible for enhanced biological phosphorus removal are not capable of denitrification, and no provisions for denitrification are necessary in modeling their behavior. However, Lotter (1985) found that a number of strains of *Acinetobacter*

were capable of denitrification; many researchers (e.g., Bortone et al., 1994; Kuba et al., 1993; Wanner et al., 1992; and Vlekke et al., 1988) have documented phosphorus uptake in the anoxic zone of biological phosphorus removal systems, indicating that the organisms responsible for biological phosphorus removal can use nitrates as an electron acceptor for the oxidation of stored carbon in the absence of free oxygen.

4. MATERIALS AND METHODS

4.1 DESCRIPTION OF THE PILOT-SCALE FGR-SGR FACILITY

The pilot-scale FGR-SGR facility was located on the grounds of B.C. Research Inc., near the pilot-scale activated sludge-type biological nutrient removal facility operated by the University of British Columbia (UBC) Department of Civil Engineering. A flow of raw sewage was available from the UBC facility for use in the FGR-SGR system. Completely mixed storage tanks sufficient for 36 hours independent operation of the FGR-SGR facility were provided, to ensure an adequate daily supply of sewage. A side-stream primary sludge fermentation process was incorporated into the design, to provide the soluble carbonaceous substrates necessary for enhanced biological phosphorus removal.

The pilot plant was designed to treat domestic wastewater in two parallel, identical process trains. Each train was sized for a raw sewage influent flow of approximately 7 m^3/d (4.8 L/min). The process design was developed from the design of the full-scale facility at Salmon Arm. The Salmon Arm process included endogenous denitrification of the return settled biosolids from the secondary clarifier underflow, and fermentation of settled solids in the primary clarifier using the activated primary concept described by Barnard(1984). In contrast, the pilot plant included denitrification using wastewater BOD as a carbon source, and fermentation of primary solids in a dedicated reactor.

A schematic of a single pilot-scale FGR-SGR process train is shown in Figure 1. Each process train included complete-mix tankage for fermentation of primary solids, and anaerobic (neither oxygen or nitrates present), anoxic (nitrates but no oxygen

present), and aerobic (dissolved oxygen present) treatment of the process mixed liquor, as well as two fixed growth reactors (FGRs), and a primary and secondary clarifier. The suspended growth reactors (SGRs) in each train were contained within a single rectangular tank, with a total volume of approximately 4,200 L. Moveable sludge-tight baffles separating the SGRs allowed controlled changes in the sizes of the individual SGR basins. All process flow rates, including the irrigation rate of the FGRs, were controlled by variable speed pumps.
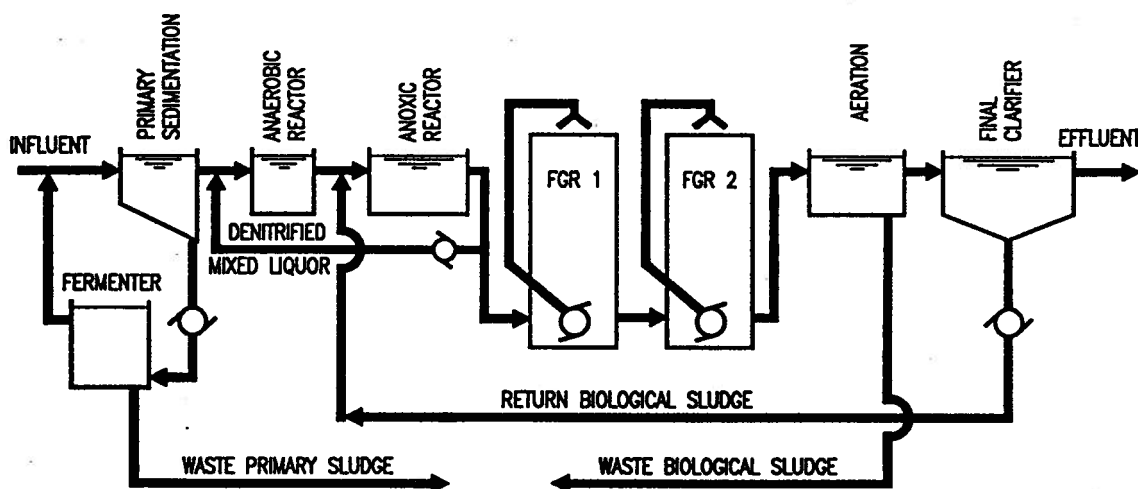
Figure 1 - Schematic of the FGR-SGR Pilot Plant

Each FGR cell had a cross sectional area of 610 mm x 610 mm, and a height of 5.5 m. The FGR media was 60^0 crossflow plastic corrugated sheets, with a specific surface area of 88 m^2/m^3. To minimize edge effects, 7.5 cm-wide flow deflectors were attached to the cell walls at 610 mm intervals and inserted between the media modules, to direct the flow back into the media, and reduce streaming down the cell walls. Sampling ports were located at 1.2 m intervals throughout the depth of each

FGR cell. At each sampling port, a 3 cm diameter hole penetrating to the center of the media allowed the collection of a sample across the media cross-section. To take a sample, an open-top collection tube was inserted into the hole to the center of the media cross-section, and process liquid collected by the tube was directed into a filter funnel.

As described in the literature review, most of the available design information for nitrification in trickling filters is based on tertiary systems, where the media is irrigated with secondary clarifier overflow. The FGR-SGR process more closely resembles a single-stage system, where the media is irrigated with primary clarifier overflow. In a study of single-stage systems in the United States, USEPA (1991) reported hydraulic loadings in the range 0.13 - 0.24 L/s/m², total BOD loadings in the range 1.17-1.91 g/m²/d (soluble BOD loadings not reported), and ammonia loadings in the range 0.19-0.28 g N/m²/d. The design hydraulic loading for the pilot-scale FGR was in the range 0.6-1.1 L/s/m², to ensure adequate cascade aeration of the activated sludge, a rate which is much higher than the ranges reported for single-stage systems, and which is typical of tertiary systems (e.g., USEPA, 1993). The soluble BOD loading for the pilot-scale FGR was 0.65 g/m²/d, based on the total media surface .area of each process train; well under the recommended maximum total BOD loading of 1.5 g/m²/d for single-stage systems (USEPA, 1991). The ammonia loading based on the total media area of each process train was approximately 0.22 g N/m²/d, within the reported range for single-stage systems.

The anaerobic-anoxic sequence (Figure 1) was modeled on a variation of the activated sludge University of Cape Town (UCT) process. The reason for selecting a UCT-type configuration for the unaerated reactors was that the UCT process can be designed to ensure a zero discharge of nitrates in the return sludge stream entering the anaerobic basin (e.g., Wentzel et al., 1991a and Pitman, 1991); as described in the literature

review, the presence of nitrates in the anaerobic reactor is known to suppress enhanced biological phosphorus removal.

The flow of settled biosolids from the final clarifier was returned to the anoxic reactor, where it mixed with the flow leaving the anaerobic reactor (Figure 1). In the anoxic reactor, carbonaceous substrates (BOD) remaining in the process liquid leaving the anaerobic reactor could be utilized for bacterial denitrification of nitrates in the return sludge.

Denitrified mixed liquor from the anoxic reactor was pumped to the anaerobic reactor, where it mixed with the primary clarifier overflow for carbon storage-phosphorus release (Figure 1). Volatile fatty acids present in the raw sewage feed were supplemented by fermentation of settled primary solids. Settled primary solids were pumped to the fermenter, and the fermenter overflow was returned to mix with the raw influent to the primary clarifier (Figure 1). The soluble products of fermentation were carried out with the clarifier overflow, as the fermented solids resettled along with the influent primary solids. The fermenter design was developed from work described by Rabinowitz and Oldham (1985).

In activated sludge biological nutrient removal systems (e.g., the UCT process), the effluent from the anaerobic-anoxic sequence flows to an aerated basin, where phosphorus uptake and nitrification of ammonia are accomplished by suspended process bacteria, using oxygen as an electron acceptor (e.g., Wentzel et al., 1991a and 1991b).

In the FGR-SGR pilot plant, the mixed liquor from the anoxic reactor was irrigated over two trickling filters in series (FGR 1 and FGR 2 in Figure 1); dissolved oxygen was provided by cascade aeration, and oxidation of residual carbonaceous substrates in the process liquid and nitrification of ammonia could be accomplished by fixed

growth organisms attached to the FGR media.

The aeration reactor downstream of the FGRs was designed to function as a polishing step, for flocculation of sloughed FGR solids and bacterial removal of residual BOD, ammonia, and phosphorus. Mixed liquor from the aeration reactor flowed to the final clarifier, where the clarified effluent was discharged to the sewer, and settled biological solids were returned to the anoxic reactor.

Designated volumes of primary solids and secondary biosolids were wasted daily from the fermenters and the aeration basins, respectively.

4.2 MONITORING OF PILOT PLANT PERFORMANCE

Overall process performance was monitored by the analysis of 24-hour composite samples of process influent (primary clarifier overflow) and process effluent (secondary clarifier overflow), for five-day biochemical oxygen demand (BOD_5), ammonia nitrogen (NH_4^+), nitrate+nitrite nitrogen (NO_x), total kjeldahl nitrogen (TKN), total phosphorus, orthophosphate (PO_4^{3-}), and total suspended solids (TSS). Grab samples taken from designated points in the process trains were analyzed for the above parameters, as well as for volatile fatty acids (VFA), so that relevant mass balances on individual unit operations within the process could be carried out. Samples for TSS taken from the aeration basins were also analyzed for volatile suspended solids (VSS) and percent phosphorus by dry weight. Sampling and analysis of the above parameters was conducted twice weekly. A summary of sample locations, sampling frequencies, and sample analyses is given in Table 1. The analyses shown in Table 1 were used to conduct mass balances on individual reactors within the process for relevant parameters. A summary of the mass balances is given in Table 2; all mass balances were normalized to reflect nominal HRT (reactor volume divided by the process influent flow rate), rather than actual HRT (reactor volume

divided by process influent flow rate plus recycle flow rates). The method of calculating the mass balances for each reactor is shown in Appendix 1. Grab samples were taken periodically from the sampling ports in the FGR cells, to assess the rates of phosphorus removal and nitrification across the media profile.

Table 1 - Summary of Sampling and Analysis Parameters

Analysis	Sample Location (Both A and B Sides)	Frequency (weekly, per side)	Total Samples (weekly)
Total Suspended Solids (TSS)	Raw Plant Influent	2	4
	Primary Clarifier Overflow	2	4
	Fermenter Mixed Liquor	2	4
	Anaerobic Mixed Liquor	2	4
	Anoxic Mixed Liquor	2	4
	Aeration Mixed Liquor	2	4
	Return Biological Sludge	2	4
	Final Clarifier Overflow	2	4
	Total	16	32
Total Biochemical Oxygen Demand (Total BOD$_5$)	Raw Plant Influent	1	2
	Primary Clarifier Overflow	1	2
	Final Clarifier Overflow	1	2
	Total	3	6
Filtered Biochemical Oxygen Demand (Filtered BOD$_5$)	Raw Plant Influent	1	2
	Primary Clarifier Overflow	1	2
	Anaerobic Zone Effluent	1	2
	Anoxic Zone Effluent	1	2
	FGR Cell 1 Effluent	1	2
	FGR Cell 2 Effluent	1	2
	Final Clarifier Overflow	1	2
	Return Biological Sludge	1	2
	Total	8	16
Total Volatile Fatty Acids (VFAs)	Raw Plant Influent	2	4
	Fermenter Overflow	2	4
	Primary Clarifier Overflow	2	4
	Anaerobic Zone Effluent	2	4
	Anoxic Zone Effluent	2	4
	Total	10	20

Table 1 - Summary of Sampling and Analysis Parameters (cont.)

Analysis	Sample Location (Both A and B Sides)	Frequency (weekly, per side)	Total Samples (weekly)
Total Phosphorus	Raw Plant Influent	2	4
	Primary Clarifier Overflow	2	4
	Final Clarifier Overflow	2	4
	Dried Biological Sludge	2	4
	Total	8	16
	Primary Clarifier Overflow	2	4
Orthophosphate	Anaerobic Zone Effluent	2	4
(PO_4^{3-}) and	Anoxic Zone Effluent	2	4
Nitrite + Nitrate	FGR Cell 1 Effluent	2	4
(NO_x)	FGR Cell 2 Effluent	2	4
	Aeration Zone Effluent	2	4
	Final Clarifier Overflow	2	4
	Return Biological Sludge	2	4
	Total	16	32
Total Kjeldahl	Raw Plant Influent	2	4
Nitrogen (TKN)	Primary Clarifier Overflow	2	4
	Final Clarifier Overflow	2	4
	Total	6	12
Ammonia	Raw Plant Influent	2	4
Nitrogen	Primary Clarifier Overflow	2	4
(NH_4^+)	Anoxic Zone Effluent	2	4
	FGR Cell 1 Effluent	2	4
	FGR Cell 2 Effluent	2	4
	Final Clarifier Overflow	2	4
	Total	12	24

Table 2 - Summary of Reactor Mass Balances

Process Reactor	Mass Balance Parameters
Primary Clarifier/Fermenter Loop	Total Volatile Fatty Acid Production
	Ammonia Production
	Filtered BOD_5 Production
Anaerobic Reactor	Total Volatile Fatty Acid Removal
	Orthophosphate Release
	Filtered BOD_5 Removal
Anoxic Reactor	Orthophosphate Release/Uptake
	Filtered BOD_5 Removal
	Nitrate/Nitrite Removal
Fixed Growth Reactor (FGR) Cell 1	Orthophosphate Uptake
	Ammonia Removal
	Nitrate/Nitrite Production
	Filtered BOD_5 Removal
Fixed Growth Reactor (FGR) Cell 2	Orthophosphate Uptake
	Ammonia Removal
	Nitrate/Nitrite Production
	Filtered BOD_5 Removal
Aeration Reactor	Orthophosphate Uptake
	Ammonia Removal
	Nitrate/Nitrite Production
	Filtered BOD_5 Removal
Final Clarifier	Orthophosphate Release/Uptake
FGR-SGR Process Effluent	Total Phosphorus Removal
(Primary Clarifier Overflow to Final Effluent)	Total Nitrogen Removal
	Ammonia Removal
	Filtered BOD_5 Removal
Total Plant	Total Phosphorus Removal
	Total Nitrogen Removal
	Ammonia Removal
	Total BOD_5 Removal

Continuous on-line monitoring of dissolved oxygen concentration and pH was also carried out at appropriate points in the process trains. The dissolved oxygen in the aeration basins was always maintained at greater than 2 mg/L, and sodium bicarbonate was added to maintain a pH of approximately 7 in the process liquid.

4.3 BENCH-SCALE BATCH TESTS

In addition to monitoring pilot-scale process performance, bench-scale batch tests designed to simulate the performance of unit operations within the process were periodically conducted. During the earlier full-scale work at Salmon Arm, batch tests adapted from the procedure described by Comeau (1984) were found to be a valuable operational tool in gaining insight into biochemical reaction rates in the process liquid (Gibb et al., 1989).

The batch tests were designed to simulate actual flow rates and HRTs in the pilot-scale suspended growth reactors. For example, in the pilot plant anaerobic reactor, the denitrified recycle flow from the anoxic reactor mixed with the primary clarifier overflow in a ratio of 2:1. For a typical batch test, grab samples of mixed liquor from the anoxic reactor and the primary clarifier overflow were mixed together at time=0 in a ratio of 2:1 in a 2.8 L batch reactor, which was then held under fully mixed, anaerobic conditions for a duration which matched the actual HRT of the process anaerobic reactor. Samples were withdrawn regularly from the batch reactor and analyzed for the appropriate parameters.

In the pilot plant anoxic reactor, the return settled biosolids mixed with the flow from the anaerobic reactor in a ratio of 1:3 (the return biosolids being the flow from the secondary clarifier to the anoxic reactor - see Figure 1). For a typical batch test, at the end of the anaerobic phase, an aliquot of return settled biosolids was added to the mixed liquor remaining in the batch reactor in a ratio of 1:3. The batch reactor was

kept under fully mixed, unaerated conditions for a time equal to the actual HRT of the process anoxic reactor, and the mixed liquor was then aerated for a time equal to the actual HRT of the process suspended growth aerobic reactors (i.e., the actual HRT in the FGR catchbasins, effluent sump, and aeration reactor). Again, samples were withdrawn at regular intervals and analyzed for the appropriate parameters.

The batch tests described above simulated a plug flow situation; that is, the bulk solution substrate concentration diminished (or increased) over time. The pilot plant reactors, on the other hand, were completely mixed flow-through basins, with (pseudo) steady-state substrate concentrations. Therefore, the batch test results could be expected to reflect conditions in the pilot plant reactors only when the substrate reaction rate was observed to be zero order with respect to concentration; that is, when the rate of substrate uptake (or release) did not change with decreasing (or increasing) bulk solution substrate concentration.

The batch tests were used to simulate the performance of the suspended growth reactors only; due to practical difficulties, FGR performance could not be simulated in bench-scale batch tests. Changes to the design and operational modes of the two pilot-scale treatment trains were made in light of the batch test results, to explore the long-term effects of those changes. Batch tests were further used to conduct "what if" scenarios, to evaluate the short-term effects of controlled changes in process internal recycle flow rates.

4.4 SAMPLE PRESERVATION AND ANALYSIS

4.4.1 Biochemical Oxygen Demand

Samples for total five-day biochemical oxygen demand were analyzed immediately after collection according to APHA et al. (1992), using an Orion 97-08-00 dissolved oxygen membrane electrode. Samples for soluble five-day biochemical oxygen

demand were filtered through Whatman No. 4 filter papers and analyzed as above. Hach Nitrification Inhibitor Formula 2533 was used to eliminate nitrogenous oxygen demand.

4.4.2 Total Suspended Solids and Volatile Suspended Solids

Samples for total suspended solids and volatile suspended solids were analyzed immediately after collection, according to APHA et al. (1992).

4.4.3 Total Phosphorus

Liquid samples for total phosphorus were frozen immediately after collection. Samples of process suspended solids to be analyzed for percent phosphorus by weight were separated from the liquid immediately after collection by filtering through Whatman glass fibre filters. The solid residue was oven-dried at 104°C, and finely ground for analysis. Analysis for total phosphorus was conducted on a Lachat QuikChem Automated Ion Analyzer according to QuikChem Method No. 10-115-01-1-C.

4.4.4 Total Kjeldahl Nitrogen

Samples for total kjeldahl nitrogen were frozen immediately after collection. Samples for soluble total kjeldahl nitrogen were filtered through Whatman No. 4 filter papers and then frozen. Analysis was conducted on a Lachat QuikChem Automated Ion Analyzer according to QuikChem Method No. 10-107-06-2-E.

4.4.5 Orthophosphate

Samples for orthophosphate were filtered through Whatman No. 4 filter papers immediately after collection, preserved by adding 1 drop per 10 mL sample of phenyl mercuric acetate solution (0.1 g phenyl mercuric acetate in 20 mL acetone and 80 mL distilled water), and stored at 4° C for up to one week. Samples were analyzed on a

Lachat QuikChem Automated Ion Analyzer according to QuikChem Method No. 10-115-01-1-D.

4.4.6   Nitrate and Nitrite Nitrogen

Samples for nitrate and nitrite nitrogen were filtered through Whatman No. 4 filter papers immediately after collection, preserved by adding 1 drop per 10 mL sample of phenyl mercuric acetate solution (0.1 g phenyl mercuric acetate in 20 mL acetone and 80 mL distilled water), and stored at 4^0 C for up to one week. Samples were analyzed on a Lachat QuikChem Automated Ion Analyzer according to QuikChem Method No. 10-107-04-1-E, a method which gives only the total sum of nitrate plus nitrite, and does not distinguish between the two.

4.4.7   Ammonia Nitrogen

Samples for ammonia nitrogen were filtered through Whatman No. 4 filter papers immediately after collection, preserved by reducing the pH of the sample to less than 2 through the addition of sulfuric acid, and stored at 4^0 C for up to one week. Samples were analyzed on a Lachat QuikChem Automated Ion Analyzer according to QuikChem Method No. 10-107-06-1-D.

4.4.8   Volatile Fatty Acids

Samples for volatile fatty acids were filtered immediately after collection using Whatman No. 4 filter papers, preserved by adding 0.1 mL of 2% phosphoric acid solution to 1 mL sample, and stored at 4^0 C for up to one week. Analyses for acetic acid, propionic acid, butyric acid, and iso-butyric acid were conducted according to Supelco GC Bulletin 751G, using a Hewlett Packard 5880A Series Gas Chromatograph.

4.4.9 Dissolved Oxygen

Dissolved oxygen in the aeration basins was monitored on-line using Rosemount Model 499 DO Sensor hooked up to Rosemount Analytical Microprocessor Analyzer Model 1054A.

4.4.10 pH

The pH of the fermenters and the anaerobic, anoxic, and aerobic suspended growth reactors was monitored on-line using Rosemount pH/ORP Sensor Model 399-07 hooked up to Rosemount Analytical pH Analyzer Model 1054ApH.

4.4.11 Temperature

The temperature of the fermenters and the anaerobic, anoxic, and aerobic suspended growth reactors was monitored using temperature sensors which were built into the Rosemount Analytical Probes described above for dissolved oxygen and pH.

4.5 EXPERIMENTAL DESIGN

4.5.1 Pilot Plant Design and Operating Parameters

Operation of the two pilot-scale FGR-SGR process trains treating a common waste stream allowed parallel comparisons between different design and operational modes. After both trains were operated in an identical mode for an acclimation period, a change was made to the experimental train. The performance of the experimental train was then compared to the control train over a designated period, using the *t* test for paired comparisons in Microsoft Excel Version 4.0.

The *t* test for paired comparisons is designed to improve the precision of comparison between the means of two data sets, by making comparisons within matched pairs of experimental data, where the experimental material may be subject to uncontrolled variations between different matched pairs, but each matched pair is tested on a

common sample of experimental material. Inferences are made about the difference between two means by making inferences about the mean of the differences between the matched pairs (Montgomery, 1984).

The null hypothesis in the paired comparison test is that there is no difference between the two means. The test statistic t_0 is calculated by dividing the mean of the differences between the matched pairs by the product of the standard deviation of the differences and the square root of the number of matched pairs (n). The value of t_0 is compared to the t distribution with n-1 degrees of freedom at a pre-selected level of significance. The value of the t distribution may be based on the one-tailed test or the two-tailed test. In both cases, the null hypothesis is that the means of the two data sets are equal. The one-tailed test is used where the alternative to the null hypothesis is that a designated sample set (say sample set A) will produce a mean which is greater than the mean of the other sample set (say sample set B); in this case, if the value of t_0 is greater than the appropriate value of the t distribution, the null hypothesis is rejected, and it may be concluded that the mean of sample A is significantly greater than that of sample B. The two-tailed test is used where the alternative hypothesis must allow for the designated sample set to produce a mean which might be either greater than or less than the mean of the other sample set; in this case, if the absolute value of t_0 is greater than the appropriate value of the t distribution, the null hypothesis is rejected, and it may be concluded that the mean of sample A is significantly different from that of sample B (Walpole, 1982). Unless otherwise stated, the experimental comparisons in this study were designed using the two-tailed test, to allow for a change to the experimental process train resulting in either an improvement or a deterioration in process performance, compared to the control process train. Note that since the standard deviation is in the denominator when calculating the value of t_0, a relatively high standard deviation in the differences between the matched pairs will reduce the value of t_0, and thereby reduce the chance

of detecting a significant difference between the two means.

For this study, the matched pairs were the samples taken from the A and B Sides at the same location in the process and on the same day. As described above, the *t* test for paired comparisons is designed to factor out differences in the experimental material which are common to both members of each matched pair. In this case, the experimental material was the process influent, which could be expected to vary in character and strength on a daily and seasonal basis. Further, environmental conditions such as temperature could be expected to vary daily and seasonally. The experimental treatment was the single controlled difference in the design or operating parameter between the A and B Sides. Since the A and B Side processes were operated in parallel treating a common influent sewage, both sides were subject to the same daily and seasonal variations in influent quality and strength, and to the same uncontrolled variations in environmental conditions (e.g., temperature). The *t* test comparisons were carried out at the 0.05 level of significance. That is, there was a 5% probability that the null hypothesis was rejected in error.

A summary of the SGR reactor sizes and nominal and actual HRTs for each of the five study phases is given in Table 3. The nominal HRT is defined as the reactor volume divided by the process influent flow rate. The actual HRT is defined as the reactor volume divided by the sum of the process influent flow rate plus any internal recycle streams influent to that reactor. The initial sizes for the anaerobic, anoxic, and aeration basins were based on the results of previous work at the full-scale FGR-SGR plant at Salmon Arm, and on the HRTs in the adjacent UCT-type activated sludge pilot plant operated by the UBC Department of Civil Engineering. Reductions in the HRTs of designated reactors were based on the results of process monitoring and batch testing, and were governed to some extent by the locations of the moveable baffles.

Table 3 - Summary of Pilot Plant Design and Operating Parameters

Phase	Objective	Design Parameters						
Phases 1-5 Aug./92 to Aug/94		Plant Influent Flow (Q)=4.8 L/min Denitrified Recycle=2Q Return Settled Biosolids Flow=Q FGR 1 Recycle Flow=FGR 2 Recycle Flow=5Q[1]						
		Reactor	Volume (L)		Nominal HRT (hr)[2]		Actual HRT (min)[3]	
			A Side	B Side	A Side	B Side	A Side	B Side
Phase 1 Mar 8/93 to May 4/93	Anoxic Reactor Optimization	Anaerobic	630	630	2.2	2.2	45	45
		Anoxic	1260	630	4.4	2.2	65	35
		Aerobic[4]	1520	1520	5.3	5.3	160	160
		Total	3410	2780	11.9	9.7		
Phase 2 Jun. 6/93 to Jul. 30/93	Anaerobic Reactor Optimization	Anaerobic	630	380	2.2	1.3	45	25
		Anoxic	630	630	2.2	2.2	35	35
		Aerobic[4]	1520	1520	5.3	5.3	160	160
		Total	2780	2530	9.7	8.8		
Phase 3 Nov. 30/93 to Feb. 23/94	Anaerobic Reactor Optimization	Anaerobic	125	380	0.4	1.3	8	25
		Anoxic	630	630	2.2	2.2	35	35
		Aerobic[4]	1520	1520	5.3	5.3	160	160
		Total	2275	2530	7.9	8.8		
Phase 4a Apr. 29/94 to Jun. 10/94	FGR Recycle Optimization	Anaerobic	125	125	0.4	0.4	8	8
		Anoxic	630	630	2.2	2.2	35	35
		Aerobic[4]	1520	1520	5.3	5.3	160	160
		Total	2275	2275	7.9	7.9		
Phase 4b Jun. 10/94 to Jul. 25/94	FGR Recycle Optimization	Anaerobic	380	380	1.3	1.3	25	25
		Anoxic	630	630	2.2	2.2	35	35
		Aerobic[4]	1520	1520	5.3	5.3	160	160
		Total	2530	2530	8.8	8.8		
Phase 5 Jul. 26/94 to Aug. 30/94	Sludge Wasting Rate Optimization	Anaerobic	380	380	1.3	1.3	25	25
		Anoxic	630	630	2.2	2.2	35	35
		Aerobic[4]	1520	1520	5.3	5.3	160	160
		Total	2530	2530	8.8	8.8		

[1] Except for Phases 4a and 4b when B Side FGR 1 Recycle Flow=FGR 2 Recycle Flow=3Q

[2] Reactor volume/process influent flow rate

[3] Reactor volume/(process influent flow rate +recycle flow rate)

[4] Includes FGR Cells 1 and 2 catchbasins @ 210 L each, FGR effluent sump @ 160L, and aeration basin @ 940 L

The results of process monitoring and batch testing during the initial acclimation phase, when both sides were operated in an identical fashion (August, 1992 through February, 1993), were used to select the first reduction in HRT. For the first HRT reduction, the volume of the anoxic basin on the B side process train was halved on March 8, 1993, reducing the anoxic actual HRT from 65 minutes to 35 minutes. The HRT of the anoxic reactor on the A side was held at 65 minutes (Phase 1 in Table 3). All other operating parameters between the two sides were held at the original values. When steady-state operation on both sides was established, the performance of the two process trains was compared, through process monitoring and bench-scale batch tests. Following the satisfactory completion of Phase 1, the volume of the A Side anoxic reactor was reduced to match that of the B Side, and both sides were operated identically through another acclimation phase.

During Phase 1, batch test investigations indicated that the anaerobic volume in the pilot plant could also be substantially reduced. After the acclimation period following Phase 1, the volume of the anaerobic reactor on the B Side process was reduced by 40%, cutting the actual HRT from 45 minutes to 25 minutes. The A Side anaerobic HRT was held at 45 minutes (Phase 2 in Table 3). Again, steady-state operation was established, and the performance of the two sides was compared through process monitoring and batch testing.

The above approach was used to assess the effects of a further reduction in actual anaerobic HRT from 25 minutes to 8 minutes on the A Side, with the B Side being held at 25 minutes (Phase 3 in Table 3).

As described later in this thesis, it was discovered during the acclimation period between Phases 2 and 3 (i.e., in late November of 1993) that periodic increases in FGR recycle rate greatly improved phosphorus removal in the pilot plant, by preventing excess solids accumulation on the FGR media. Throughout the remainder of the study

(Phases 3-5), regular hydraulic pulse loading of the FGRs was carried out on both the A and B Sides, by increasing the FGR steady-state recycle rate by a factor of four for approximately 1 hour per day.

Following the completion of Phase 3, the actual HRT of the anaerobic reactor on the B Side was reduced to 8 minutes, to match that of the A Side. The 8 minute anaerobic HRT was selected on the basis of the results of previous work to optimize reactor size (Phase 3). Both sides were then operated identically for an acclimation period until April 28, 1994, according to the parameters shown for the A Side during Phase 3 in Table 3.

Due to time and budget constraints, it was not possible to investigate the optimum HRT of the aeration basin in the pilot plant. Therefore, after the acclimation period following Phase 3, the investigation into FGR recycle rate (Phase 4) was begun. The purpose of Phase 4 was to determine the effects of FGR recycle rate on nitrification and phosphorus removal in the pilot plant. On April 29, 1994, the recycle rate on the B Side process for FGR Cells 1 and 2 was reduced from 24 L/min (five times the process influent flow rate) to 14 L/min (three times the process influent flow rate). All other parameters were maintained identical between the two sides (Table 3).

Note that Phase 4 was divided into two separate periods, designated Phase 4a and Phase 4b in Table 3. Throughout Phases 4a and 4b, the FGR recycle rate was 24 L/min on the A Side and 14 L/min on the B Side. During Phase 4a (April 29 through June 9 of 1994), the anaerobic actual HRT on both sides was 8 minutes; during Phase 4b (June 10 through July 25 of 1994), the anaerobic actual HRT on both sides was 25 minutes. A deterioration in phosphorus removal was observed on Both the A and B Sides during the acclimation period following Phase 3 and throughout most of Phase 4. It was suspected that the deterioration was caused by a change which rendered the 8 minute anaerobic HRT insufficient for good phosphorus removal. Accordingly, the

anaerobic reactors on both the A and B Sides were expanded to an actual HRT of 25 minutes in early June of 1994 (Table 3). The deterioration in phosphorus removal and changes in the operating condition of the pilot plant following Phase 3 are discussed later in this thesis.

During Phase 4, process monitoring and batch testing indicated that the performance of the A Side (FGR recycle rate=5Q) was not significantly better than that of the B Side (FGR recycle rate=3Q). Accordingly, in late July of 1994, the FGR recycle rate on the A Side was reduced to 3Q to match the B Side, and the investigation into optimum system operating MLSS concentration (Phase 5) was begun. Due to project time and budget constraints, it was not possible to allow an acclimation period between Phases 4 and 5.

The solids retention time (SRT) in an activated sludge reactor is defined as the mass of cells in the system divided by the mass of cells wasted from the system per day. The operating mixed liquor suspended solids (MLSS) concentration in the system is a function of the SRT, since wasting more cells per day generally results in a lower MLSS concentration. In fixed growth systems, the SRT is more difficult to determine. As bacteria grow and reproduce in the fixed biofilm, the film becomes thicker, and flocs of biomass are eventually sloughed off by hydraulic forces.

The FGR-SGR process incorporates both fixed and suspended growth components; therefore, a simple calculation of SRT based on the volume of the aeration basin divided by the volume of secondary sludge wasted daily does not account for the SRT of the fixed growth in the FGR, nor does it account for suspended organisms which may adhere to the fixed biofilm for an unknown duration, before being sloughed off. Therefore, for the purposes of this study, the aeration basin MLSS concentration was chosen as the operating variable for the investigation of solids retention time. That is, the daily sludge wasting rate was adjusted to keep the aeration basin MLSS

concentration as close as possible to a designated value.

During Phases 1-4, the secondary sludge wasting rate was adjusted to keep the aeration basin MLSS concentration on both the A and B Sides as close to 3000 mg/L as possible. On July 26, 1994, the sludge wasting rate on the B Side was increased in order to maintain the operating aeration basin MLSS concentration as close to 2000 mg/L as possible. Sludge wasting on the A Side was done at a rate sufficient to maintain the operating aeration basin MLSS concentration as close to 3000 mg/L as possible. All other parameters for both sides were kept identical between the two sides (see Phase 5 in Table 3).

During Phase 5, the average daily volume of mixed liquor wasted from the A Side aeration basin was 110 L, and the average daily volume wasted from the B Side was 210 L. As in Phases 1-4, the performance of the A and B Sides was compared through process monitoring and batch testing. Following the completion of Phase 5 at the end of August, 1994, the pilot plant was de-commissioned.

Sampling across the FGR profile to establish substrate removal rates with media depth on three separate occasions did not yield useful results. Due to a wide scatter in the data, no consistent step-wise removal of phosphorus or production of nitrate between adjacent sampling ports was observed across the FGR media profile in any of the cells. The time required for the collection of a sufficient quantity of sample for analysis varied with different sampling ports, and filtering of the sample generally took 3-5 minutes. It is likely that substrate removal by suspended organisms during the sample collection and filtering masked the removal accomplished by fixed growth between sampling ports. The procedure was therefore abandoned. However, removal of phosphorus and ammonia, and production of nitrate, were consistently observed where mass balances were calculated on each FGR cell as a whole. Results of the mass balance calculations are discussed later in this thesis.

The average actual retention time of the process mixed liquor in the media of each FGR cell was periodically estimated, using the method described by Tariq (1975). The method is similar to that commonly used to estimate the average actual HRT in completely mixed activated sludge reactors (i.e., the reactor liquid volume divided by the reactor flow-through rate). According to the method of Tariq, the actual liquid retention time in the media of a trickling filter (FGR) may be estimated by the following method:

- measure the steady-state irrigation flow rate to the FGR media
- shut off the FGR irrigation flow at time=0, and measure the volume of liquid which drains from the media from time=0 until all flow from the media has stopped
- the average actual HRT of the liquid in the media is equal to the volume of liquid collected divided by the steady-state irrigation flow rate.

4.5.2 Batch Test Investigation of Denitrified Recycle Flow Rate

The denitrified recycle in the pilot plant was the internal process flow which returned denitrified mixed liquor from the anoxic reactor to the anaerobic reactor (see Figure 1). Time did not allow operation of the pilot plant at several different denitrified recycle flow rates, with time for acclimation and steady-state monitoring after each change. However, a batch test simulation was developed to investigate at bench-scale the effects of varying the denitrified recycle rate for a single cycle through the plant.

The purpose of the batch test was to investigate the effects of manipulating the denitrified recycle flow rate to dampen the effects of the diurnal flow fluctuations typically experienced at a full-scale plant. That is, when the daily peak hydraulic load arrives at a treatment plant, it results in a lower actual HRT in the process bioreactor(s), since average actual HRT equals reactor volume divided by reactor flow-through rate (reactor flow-through rate being the sum of the process influent

flow plus any internal recycle flow). Besides reducing the actual HRT, an increase in process influent flow rate also tends to dilute the MLSS concentration in the process reactor, thus increasing the food to microorganism (F/M) ratio.

The actual HRT in the anaerobic reactor of the pilot plant was equal to the anaerobic volume divided by the sum of the process influent flow rate and the denitrified recycle flow rate (see Figure 1). The batch test in this case was designed to investigate the single-cycle effects of decreasing the denitrified recycle rate to increase the actual anaerobic HRT, and also to investigate the effects of increasing the recycle to increase the MLSS concentration in the anaerobic phase and attenuate the increase in F/M ratio caused by the increase in process influent flow rate.

At the time of the batch test (August 23, 1994), the A and B Side process trains were being operated according to the parameters described for Phase 5 in Table 3. That is, the steady-state process influent flow (Q) was 4.8 L/min, the denitrified recycle rate was 9.6 L/min (2Q), the return settled biosolids recycle rate was 4.8 L/min (Q), and the actual anaerobic, anoxic, and aerated retention times were 25 minutes, 35 minutes, and 160 minutes, respectively.

For the batch test, five batch reactors were operated in parallel, according to the procedure described earlier in Section 4.3.1. The batch test parameters are summarized in Table 4. The five tests were conducted simultaneously, using grab samples taken from the A Side process train only. The designated aliquot of sample for each batch reactor was taken from a common, fully mixed grab sample of process liquid taken from the appropriate process reactor or flow stream in the A Side process. The method of calculating the volume of sample added is described in Appendix 2.

The simulated process influent flow for all five reactors in the batch test was 1.65

times the average daily flow rate (Q) of 4.8 L/min. Increasing the process influent flow (primary clarifier overflow) to 1.65Q (i.e., 7.92 L/min) would result in a reduction in actual anaerobic HRT from 25 minutes to 22 minutes in the A Side process anaerobic reactor. Reactor #1 (the control reactor) was configured to simulate the conditions which would result from no manipulation of the denitrified recycle rate; that is, the simulated denitrified recycle flow was maintained at the steady state value of 2Q (9.6 L/min). Reactors #2 and #3 were configured to simulate the effects of reducing the denitrified recycle rate to 1.5Q and Q, respectively, to increase the actual anaerobic HRT. Reactors #4 and #5 were configured to simulate the effects of increasing the denitrified recycle rate to 2.5Q and 3Q, respectively, to decrease the dilution of anaerobic MLSS, thereby increasing the proportion of MLSS subjected to anaerobic conditions.

Table 4 - Summary of Parameters for Investigation of Denitrified Recycle Rate

1 Reactor #	2 Simulated Denitrified Recycle (times Q)[1]	3 Volume of Sample Added (mL)[2]			4 Actual Hydraulic Retention Time (min)[3]		
		Process Influent[4] (1.65 Q)	Anoxic Liquor[5] (column 2)	Return Biosolids[6] (Q)	Anaerobic	Anoxic	Aerobic
1	2Q	1265	1535	600	22	28	120
2	1.5Q	1465	1335	675	25	32	120
3	Q	1745	1055	765	30	36	120
4	2.5Q	1115	1685	545	19	25	120
5	3Q	995	1805	495	17	23	120

[1] Q=4.8 L/min
[2] See Appendix 2 for sample calculation
[3] Simulated Process Influent Flow=1.65Q, Simulated Return Biosolids Flow=Q
[4] From primary clarifier overflow pipe
[5] From anoxic reactor near effluent weir
[6] From secondary clarifier underflow return pipe

It should be noted that the above procedure simulated the reactor MLSS concentrations which would result only after the simulated flow rates had continued for a sustained period. In a full-scale situation, the change in reactor MLSS concentration resulting from a change in the steady-state recycle flow rate would happen gradually, and might not reach the steady-state value before the recycle rate was returned to its original value after the peak flow was over. The batch test described above was designed only to give an indication of the effects of varying the denitrified recycle rate to attenuate the effects of plant influent flow fluctuations; pilot-scale or full-scale operation over several weeks or months would be required to produce design data.

4.5.3 Batch Test Investigation of Settled Biosolids Return Flow Rate

The settled biosolids recycle in the pilot plant was the internal process flow which returned settled biosolids from the final clarifier to the anoxic reactor (see Figure 1). As described earlier for the investigation of denitrified recycle rate, time did not allow operation of the pilot plant at several different biosolids recycle flow rates, with time for acclimation and steady-state monitoring after each change. However, a batch test simulation was developed to investigate at bench-scale the single-cycle effects of varying the biosolids recycle rate.

The actual HRT in the pilot plant anoxic reactor was equal to the anoxic volume divided by the sum of the process influent flow, the denitrified recycle flow, and the settled biosolids recycle flow. The aerated HRT was equal to the aerated volume divided by the sum of the process influent flow and the settled biosolids recycle flow (see Figure 1). Similar to the investigation of denitrified recycle flow described earlier, the purpose of the batch test in this case was to investigate the effects of reducing the settled biosolids recycle rate to increase the anoxic and aerated actual HRTs, and also to investigate the effects of increasing the biosolids recycle flow to dampen the

dilution effects of the simulated peak hydraulic load on the MLSS concentration in the anoxic and aerated reactors. In a full-scale plant, reduction of the settled biosolids recycle rate could also be used to dampen hydraulic shocks to the secondary clarifier during the daily peak load.

At the time of the batch test (August 23, 1994), the A and B Side process trains were being operated according to the parameters described for Phase 5 in Table 3. That is, the steady-state process influent flow (Q) was 4.8 L/min, the denitrified recycle rate was 9.6 L/min (2Q), the return settled biosolids recycle rate was 4.8 L/min (Q), and the actual anaerobic, anoxic, and aerated retention time were 25 minutes, 35 minutes, and 160 minutes, respectively.

For the batch test, five batch reactors were operated in parallel according to the procedure described earlier in Section 4.3.1. The batch test parameters are summarized in Table 5. All aliquots of process liquid for the batch test were taken from common, completely mixed samples taken from the A Side only. The method of calculating the volume of sample added is described in Appendix 2.

Reactor #1 (control) was configured to simulate the effects of maintaining the biosolids recycle rate at the steady state value of Q (4.8 L/min). The simulated process influent flow was again 1.65 times the average daily flow rate (Q) of 4.8 L/min, for all five reactors. Increasing the process influent flow (primary clarifier overflow) to 1.65Q (i.e., 7.92 L/min) would result in a reduction in anoxic actual HRT from 35 minutes to 28 minutes, and a reduction in aerated actual HRT from 160 minutes to 120 minutes in the pilot plant; the control reactor (#1) was configured to reflect those conditions. Reactors #2 and #3 were configured to simulate the effects of reducing the settled biosolids recycle rate to 0.5Q and 0.25Q, respectively, to increase the actual anoxic and aerated HRTs. Reactors #4 and #5 were configured to simulate the effects of increasing the settled biosolids recycle rate to 1.5Q and 2Q, respectively, to

decrease the dilution of reactor MLSS.

Table 5 - Summary of Parameters for Investigation of Settled Biosolids Recycle Rate

1 Reactor #	2 Simulated Settled Biosolids Recycle (times Q)[1]	3 Volume of Sample Added (mL)[2]			4 Actual Hydraulic Retention Time (min)[3]		
		Process Influent[4] (1.65Q)	Anoxic Liquor[5] (2Q)	Return Biosolids[6] (column 2)	Anaerobic	Anoxic	Aerobic
1	Q	1265	1535	600	22	28	120
2	0.5Q	1265	1535	335	22	32	147
3	0.25Q	1265	1535	180	22	34	166
4	1.5Q	1265	1535	815	22	25	100
5	2Q	1265	1535	990	22	23	87

[1] Q=4.8 L/min
[2] See Appendix 2 for sample calculation
[3] Simulated Process Influent Flow=1.65Q, Simulated Denitrified Recycle Flow=2Q
[4] From primary clarifier overflow pipe
[5] From anoxic reactor near effluent weir
[6] From secondary clarifier underflow return pipe

As described earlier for the batch test simulation of manipulating the denitrified recycle rate, note that the above procedure simulated the reactor MLSS concentrations which would result only after the simulated flow rates had continued for a sustained period, and if the solids concentration of the secondary clarifier sludge blanket did not change. In a full-scale situation, the change in reactor MLSS concentration resulting from a change in the steady-state recycle flow rate would happen gradually, and might not reach the steady-state value before the recycle rate was returned to its original value after the peak flow was over. Further, a change in the return settled biosolids flow rate would ultimately result in a change in the MLSS concentration in the secondary clarifier sludge blanket and in the return biosolids flow stream. However, the depth and density of the final clarifier sludge blanket cannot be

estimated, and an accurate mass balance of the dynamic changes in MLSS concentration in the clarifier underflow is not possible. The batch test described above was designed only to give an indication of the effects of varying the return settled biosolids flow rate to attenuate the effects of plant influent flow fluctuations; pilot-scale or full-scale operation over several weeks or months would be required to produce design data.

5. RESULTS

5.1 PILOT PLANT PERFORMANCE

As described in Chapter 4, pilot plant performance in response to design changes was evaluated through parallel comparisons between the control and experimental process trains, using mass balances on individual reactors and over the entire process. The results of the parallel comparisons are presented in the following sections for each of the parameters studied. The data show that the concentrations of suspended solids, biochemical oxygen demand, total nitrogen, and total phosphorus in the pilot plant influent were generally 70%-80% of those typically found in a domestic sewage in British Columbia. Dilution of the influent sewage to the pilot plant was likely due to infiltration of groundwater into the local sewer system.

As described in Chapter 4, the temperature of the process liquid in the suspended growth reactors was monitored throughout the study. Temperatures in the A and B Side aeration basins were in the range 9-23 °C during Phases 1-5. No statistically significant correlations between pilot plant performance and temperature were observed for any of the parameters discussed below. Note that process phosphorus removal was most effective during the coldest period (Phase 3 - see Section 5.1.7), when aeration basin temperatures were in the range 9-15 °C.

5.1.1   Total Suspended Solids

The secondary sludge wasting rate was adjusted to maintain the aeration basin MLSS concentration on both sides as close as possible to 3,000 mg/L for Phases 1-4 (i.e., from start-up of the plant to July 26, 1994). For Phase 5 (July 26-August 30, 1994), the wasting rate on the B Side was increased to maintain the aeration basin MLSS

concentration as close as possible to 2,000 mg/L (see Chapter 4). The concentration of mixed liquor suspended solids (MLSS) in the A and B Side aeration basins throughout Phases 1-5 is summarized in Figure 2. Wide fluctuations in aeration basin MLSS concentration were observed during the period November, 1992, through November, 1993. The MLSS concentration periodically tended to decrease rapidly over a period of one to two weeks. The decreases in MLSS concentration were greater on the A Side than on the B Side, and the decreases on both sides continued even when wasting of secondary sludge was discontinued altogether. The MLSS concentration reached a minimum level, and then increased rapidly (see March, May, and October of 1993 in Figure 2).

Visual observations through the FGR sampling ports in November of 1993 indicated that a significant degree of solids accumulation on the FGR media was taking place. It is likely that suspended growth biomass tended to become attached to the FGR media over time, resulting in a gradual decrease in MLSS concentration in the suspended growth reactors; this appeared to be followed by periodic major sloughing of solids from the FGR media, possibly due to a weakening of the biofilm attachment to the media caused by anaerobic conditions deep within the biofilm. The steep increases in MLSS shown during May and October of 1993 (Figure 2) were probably caused by such sloughing events. As described above, the decreases in MLSS concentration were greater on the A Side than on the B Side, indicating that solids accumulation on the FGR media was greater on the A Side. The reason for this anomaly is unknown; both sides were physically identical, and were operated with an identical FGR irrigation rate (24 L/min) during Phases 1 and 2. Possibly the greater degree of solids accumulation on the A Side was due to some difference in the hydraulic flow pattern of the process mixed liquor on the media, which developed as solids accumulation progressed.
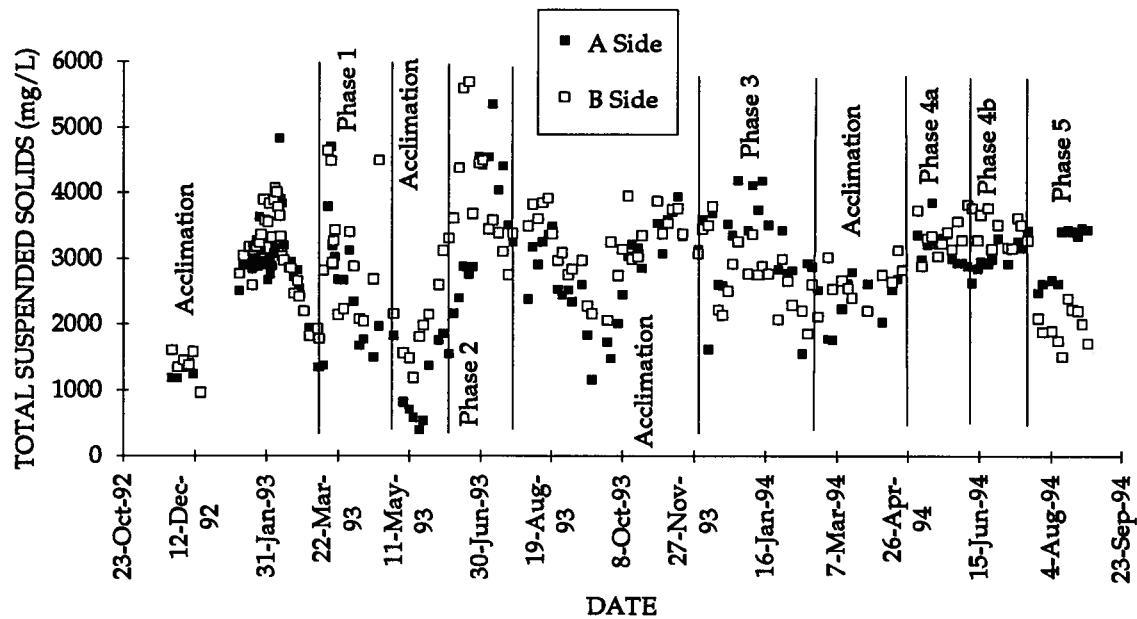
Figure 2 - Aeration Basin Mixed Liquor Suspended Solids Concentration

In order to prevent excess solids accumulation on the media, the FGR recycle (irrigation) rate on both the A and B Sides was increased fourfold for a period of two days in late November of 1993. The result was an immediate doubling in aeration basin MLSS concentration on both the A and B Sides (not shown in Figure 2), accompanied by a strong anaerobic smell. The sloughed excess MLSS were immediately wasted from the system.

Subsequently, an operational protocol was instituted during late November of 1993, where the FGR irrigation rate was increased from 24 L/min to approximately 100 L/min for a one hour period each day, to scour the media and prevent solids accumulation from becoming excessive. Following the institution of the pulse loading system, it became easier to control the process MLSS concentration, and fluctuations in MLSS concentration were greatly reduced (November, 1993, through August, 1994, in Figure 2). Visual observations through the FGR sampling ports indicated that the pulse loading prevented excess solids buildup on the media.

As discussed later in this thesis, the pulse loading regime for the FGRs had a beneficial impact on phosphorus removal. Further, the pulse loading appeared to nearly eliminate infestations of filter flies (*psychoda*) on the media. During the summer of 1993 (no pulse loading), at least two heavy infestations of filter flies occurred. During the summer of 1994 (pulse loading daily), filter flies were minimal.

The mean total suspended solids (TSS) concentrations in the process influent (primary clarifier overflow) and plant effluent (secondary clarifier overflow), for the A and B Sides throughout Phases 1-5, including the results of the *t* test comparisons between the two sides, are summarized in Table 6. During Phase 1, plant effluent mean TSS concentration was significantly higher on the A Side (24.4 mg/L) than on the B Side (15.3 mg/L), despite a significantly lower mean process influent TSS concentration on the A Side (84 mg/L), compared to the B Side (101 mg/L). The reason for the lower influent TSS concentrations on the A Side are unknown. As described above, MLSS fluctuations caused by solids accumulation on the FGR media with subsequent uncontrolled sloughing events during Phases 1 and 2 were greater on the A Side than on the B Side. It is likely that the higher effluent mean MLSS concentration observed on the A Side during Phase 1 was caused by the greater degree of solids accumulation on the FGR media on the A Side, with correspondingly greater sloughing events. Throughout Phases 2-5, plant effluent mean TSS concentration was consistently less than 15 mg/L on both sides, with no significant difference between the two, indicating that the experimental treatments carried out during Phases 2-5 had no significant effect on process effluent TSS concentration.

The mean concentrations of MLSS in the A and B Side anaerobic, anoxic, and aeration basins throughout Phases 1-5, including the results of the *t* test comparisons, are included in Table 6.

Table 6 - Results of Pilot Plant Monitoring - Total Suspended Solids

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Mean Process Influent	A Side	84	71	85	62	75
TSS Conc. (mg/L)	B Side	101	90	83	75	90
Degrees of Freedom		11	12	24	21	9
Calculated t		-2.51	-1.34	0.48	-3.12	-3.76
Critical t (alpha=0.05)		2.20	2.18	2.06	2.08	2.26
Significant Difference?		Yes	No	No	Yes	Yes
Mean Process Effluent	A Side	24.4	13.6	11.7	8.2	12.1
TSS Conc. (mg/L)	B Side	15.3	11.7	11.3	7.7	11.3
Degrees of Freedom		11	11	26	20	9
Calculated t		2.83	1.15	0.44	0.75	0.24
Critical t (alpha=0.05)		2.20	2.20	2.06	2.09	2.26
Significant Difference?		Yes	No	No	No	No
Mean Anaerobic Basin	A Side	1743	2486	2615	2215	2121
MLSS Conc. (mg/L)	B Side	1918	2879	2345	2590	1464
Degrees of Freedom		11	13	20	22	7
Calculated t		-1.18	-1.33	2.05	-2.67	12.99
Critical t (alpha=0.05)		2.20	2.16	2.09	2.07	2.36
Significant Difference?		No	No	No	Yes	Yes
Mean Anoxic Basin	A Side	2579	3654	3468	3085	3056
MLSS Conc. (mg/L)	B Side	2943	4372	3153	3380	2104
Degrees of Freedom		14	13	21	22	8
Calculated t		-1.82	-2.04	3.07	-3.51	6.67
Critical t (alpha=0.05)		2.14	2.16	2.08	2.07	2.31
Significant Difference?		No	No	Yes	Yes	Yes
Mean Aeration Basin	A Side	2421	3473	3218	3118	3089
MLSS Conc. (mg/L)	B Side	2903	3915	2922	3388	1965
Degrees of Freedom		16	13	27	21	9
Calculated t		-2.68	-1.08	2.49	-3.18	7.57
Critical t (alpha=0.05)		2.12	2.16	2.05	2.08	2.26
Significant Difference?		Yes	No	Yes	Yes	Yes

During Phases 1 and 2, the mean MLSS concentrations in the A Side anaerobic, anoxic, and aerobic SGRs were lower than that of the B Side, although according to the t test, the only significant difference at the 0.05 level was for the aeration basin during Phase 1. As described earlier, wider variation in process MLSS concentration were observed on the A Side than on the B Side during Phases 1-2 (Figure 2). The lower operating MLSS concentration on the A Side during Phases 1 and 2 was

probably caused by the greater accumulation of solids on the FGR media on the A Side.

During Phases 3-5 (November, 1993, through August, 1994), pulse loading of the FGRs allowed a much tighter control of MLSS concentration than during Phases 1 and 2 (see Figure 2). However, in spite of the tighter control and attempts to operate both sides at the same MLSS concentration, the mean anoxic and aeration basin MLSS concentrations were significantly higher on the A Side than on the B Side during Phase 3, and significantly higher on the B Side than on the A Side during Phase 4. The anaerobic basin mean MLSS concentration followed a similar pattern, although the difference between the A and B Sides was not statistically significant during Phase 3 (Table 6).

As described in Chapter 4, the aeration basin MLSS concentration was deliberately maintained at a lower value on the B Side (mean=1965 mg/L) than on the A Side (mean=3089 mg/L) during Phase 5 (Table 6). The impact of the above differences in process operating MLSS concentration on process performance will be discussed later in this thesis.

5.1.2 Biochemical Oxygen Demand

The mean concentration of total five-day biochemical oxygen demand (BOD_5) in the process influent, the mean concentration of filtered BOD_5 in the process influent, and the mean mass of total BOD_5 removed from the process influent throughout Phases 1-5, are summarized in Table 7. The t test detected no significant differences between the A and B Sides for process influent total or filtered BOD_5 concentrations, or the in process removal of total BOD_5, during any of Phases 1-5 (Table 7), indicating that the experimental treatments carried out over Phases 1-5 had no significant effect on pilot plant BOD removal. Nor did removal of BOD appear to be affected by the wide

fluctuations in suspended growth MLSS concentration during Phases 1 and 2 compared to Phases 3-5. The mean BOD_5 concentration in filtered samples of the process influent generally represented approximately 55%-60% of the mean process influent total BOD_5, except during Phase 3, when process influent filtered BOD_5 concentration was equal to approximately 70% of the total BOD_5 (Table 7).

Table 7 - Results of Pilot Plant Monitoring - Removal of Biochemical Oxygen Demand

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Mean Process Influent	A Side	97	104	112	98	114
Total BOD_5 Conc. (mg/L)	B Side	96	105	111	103	114
Degrees of Freedom		5	4	11	8	4
Calculated t		0.40	-0.98	0.39	-1.85	-0.01
Critical t (alpha=0.05)		2.57	2.78	2.20	2.31	2.78
Significant Difference?		No	No	No	No	No
Mean Process Influent	A Side	56	58	79	61	68
Filt. BOD_5 Conc. (mg/L)	B Side	54	57	78	62	60
Degrees of Freedom		5	4	8	8	4
Calculated t		1.43	0.61	0.70	-0.46	1.56
Critical t (alpha=0.05)		2.57	2.78	2.31	2.31	2.78
Significant Difference?		No	No	No	No	No
Mean Process Total BOD_5	A Side	90	99	104	93	107
Removal (mg/L)	B Side	91	101	104	97	107
Degrees of Freedom		5	4	11	7	4
Calculated t		-0.20	-2.21	-0.02	-1.32	0.00
Critical t (alpha=0.05)		2.57	2.78	2.20	2.36	2.78
Significant Difference?		No	No	No	No	No

The degree of uptake of BOD_5 in the anaerobic reactor has implications for biological phosphorus removal, since BOD uptake is an indicator of the degree of carbon storage by phosphorus-accumulating bacteria under anaerobic conditions (a more accurate indicator of anaerobic carbon storage is the removal of total volatile fatty acids, which will be discussed later in this thesis). Further, the degree of BOD removal in the anaerobic and anoxic reactors has implications for nitrification in the FGRs, since high BOD loadings are known to suppress nitrification.

The results of mass balance calculations on the removal of filtered BOD_5 in the individual process reactors, including the t test comparisons between the A and B Sides, are summarized in Table 8. The BOD_5 conversions for all of the process reactors shown in Table 8 were normalized to show mg/L of filtered BOD_5 removed per L of process influent; that is, the removals shown were based on nominal HRT, and not on actual HRT. The method of normalizing reactor substrate removals is described in Appendix 1.

As shown in Table 8, mean uptake of filtered BOD_5 in the anaerobic reactor was not significantly different between the A and B Sides during Phase 1, when the anaerobic actual HRT on both sides was 45 minutes. During Phase 2, when the anaerobic HRT on the A Side was 45 minutes compared to 25 minutes on the B Side, the mean uptake of filtered BOD_5 in the A Side anaerobic reactor was 60 mg/L, compared to a mean of only 41 mg/L on the B Side. However, the difference was not statistically significant according to the t test, possible due to a relatively large variance in the data, and the small number of samples (four). During Phase 3, when the anaerobic actual HRT on the A Side was 8 minutes compared to 25 minutes on the B Side, mean anaerobic uptake of filtered BOD_5 was significantly lower on the A Side (37 mg/L) than on the B Side (43 mg/L). During Phase 4, both the A and B Sides were operated with the same anaerobic actual HRT (8 minutes during Phase 4a and 25 minutes during Phase 4b), and BOD uptake was not significantly different between the two. During Phase 5, when both sides were operated with the same anaerobic actual HRT (25 minutes), but the A Side had a higher anaerobic mean MLSS concentration compared to the B Side (2121 mg/L vs. 1464 mg/L - see Table 6), mean BOD uptake in the A Side anaerobic reactor (42 mg/L) was significantly higher than that of the B Side (18 mg/L).

In the anoxic reactor, the conversion of filtered BOD_5 through denitrification did not differ significantly between the two sides during Phase 1, despite an anoxic actual

HRT of 65 minutes on the A Side compared to 35 minutes on the B Side (as described later in this thesis, the 35 minute anoxic HRT was sufficient for complete denitrification of available nitrate). Throughout Phases 2-5, when the anoxic actual HRT was 35 minutes on both sides, there was no significant difference in anoxic conversion of filtered BOD_5 between the two sides.

Table 8 - Results of Pilot Plant Monitoring - Filtered Biochemical Oxygen Demand

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Mean Anaerobic Filt BOD_5	A Side	20	60	37	24	42
Uptake (mg/L Influent)	B Side	21	41	43	31	18
Degrees of Freedom		4	3	10	7	3
Calculated t		-0.28	1.29	-2.84	-1.06	5.39
Critical t (alpha=0.05)		2.78	3.18	2.23	2.36	3.18
Significant Difference?		No	No	Yes	No	Yes
Mean Anoxic Filt BOD_5	A Side	29	11	15	9	5
Conversion (mg/L Influent)	B Side	26	18	11	11	7
Degrees of Freedom		4	3	11	8	3
Calculated t		0.53	-0.77	0.72	-0.35	-0.21
Critical t (alpha=0.05)		2.78	3.18	2.20	2.31	3.18
Significant Difference?		No	No	No	No	No
Mean FGR 1 Filt BOD_5	A Side	4	5	5	11	4
Conversion (mg/L Influent)	B Side	3	5	5	5	18
Degrees of Freedom		4	3	11	8	3
Calculated t		0.50	-0.11	0.09	1.32	-1.98
Critical t (alpha=0.05)		2.78	3.18	2.20	2.31	3.18
Significant Difference?		No	No	No	No	No
Mean FGR 2 Filt BOD_5	A Side	1	1	4	10	19
Conversion (mg/L Influent)	B Side	3	5	4	8	15
Degrees of Freedom		4	3	11	8	4
Calculated t		-0.62	-2.07	-0.22	0.58	1.23
Critical t (alpha=0.05)		2.78	3.18	2.20	2.31	2.78
Significant Difference?		No	No	No	No	No
Mean Aeration Basin Filt	A Side	8.5	3.8	15	9	8
BOD_5 Conv (mg/L Influent)	B Side	8.2	2.3	13	10	11
Degrees of Freedom		4	3	10	8	4
Calculated t		0.22	2.58	0.89	-0.95	-1.53
Critical t (alpha=0.05)		2.78	3.18	2.23	2.31	2.78
Significant Difference?		No	No	No	No	No

As described in the literature review, soluble BOD_5 concentrations greater than 20 mg/L in the influent to trickling filters have been found to suppress nitrification by the fixed growth. The mean concentration of filtered BOD_5 in the influent to FGR Cell 1 was in the range 6-19 mg/L throughout Phases 1-5 (Table 8). Therefore, fixed growth nitrification in the pilot plant FGRs should not have been suppressed by competition from fixed-growth heterotrophs, which used soluble BOD in the process liquid as substrate. As shown in Table 8, the mean removal of filtered BOD_5 in FGR Cells 1 and 2 and in the aeration basin did not differ significantly between the A and B Sides during any of Phases 1-5.

During Phases 1-5, the mean conversion of filtered BOD_5 across the unaerated reactors (anaerobic+anoxic) taken as a unit represented 45%-60% of the process mean removal of total BOD_5 on both the A and B Sides (i.e., an amount approximately equal to the process influent filtered BOD_5), except for the A Side during Phase 2 (72%), and the B Side during Phase 5 (25%). Conversion of the remaining (particulate) portion of the process influent total BOD_5 was likely due to hydrolysis and conversion in the unaerated reactors, and oxidation of residual degradable solids by the fixed growth on the FGR media.

In summary, the experimental treatments carried out during Phases 1-5 had no significant effect on the removal of total BOD_5 across the entire process, or on the conversion of filtered BOD_5 in any of the process reactors except the anaerobic reactor. In the anaerobic reactor, the 8 minute actual HRT resulted in a significantly lower removal of filtered BOD_5 than the 25 minute anaerobic HRT, and the operating mean anaerobic MLSS concentration of 2121 mg/L (A Side) resulted in a significantly higher removal of filtered BOD_5 than the operating mean anaerobic MLSS concentration of 1464 mg/L (B Side).

5.1.3 Total Kjeldahl Nitrogen

The mean concentration of total kjeldahl nitrogen (TKN) in the process influent and in filtered and unfiltered samples of the process effluent for the A and B Sides, including the mean process mass removal of TKN and the *t* test comparisons between the A and B Sides, are summarized in Table 9.

Table 9 - Monitoring of Pilot Plant Performance -Total Kjeldahl Nitrogen

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Mean Process Influent	A Side	19.03	23.67	21.51	20.81	23.20
Total TKN Conc. (mg N/L)	B Side	19.33	24.13	21.19	22.27	22.33
Degrees of Freedom		11	13	23	21	8
Calculated t		-0.37	-0.40	0.36	-2.75	0.61
Critical t (alpha=0.05)		2.20	2.16	2.07	2.08	2.31
Significant Difference?		No	No	No	Yes	No
Mean Process Effluent	A Side	2.47	5.14	1.96	3.05	2.96
Unfilt TKN Conc. (mg N/L)	B Side	1.98	3.13	2.05	2.69	3.46
Degrees of Freedom		11	13	24	21	8
Calculated t		4.03	1.24	-0.30	1.45	-0.30
Critical t (alpha=0.05)		2.20	2.16	2.06	2.08	2.31
Significant Difference?		Yes	No	No	No	No
Mean Process Effluent	A Side	1.46	2.76	2.50	2.38	1.90
Filt TKN Conc. (mg N/L)	B Side	1.58	2.41	1.74	1.98	2.97
Degrees of Freedom		10	9	24	21	8
Calculated t		-0.62	0.43	1.90	1.04	-0.86
Critical t (alpha=0.05)		2.23	2.26	2.06	2.08	2.31
Significant Difference?		No	No	No	No	No
Mean TKN Removal Based	A Side	17.65	21.26	18.14	18.43	21.30
on Filt Samples (mg N/L)	B Side	17.99	22.16	18.60	20.29	19.37
Degrees of Freedom		10	9	24	21	8
Calculated t		-0.44	-0.73	-0.47	-2.98	0.81
Critical t (alpha=0.05)		2.23	2.26	2.06	2.08	2.31
Significant Difference?		No	No	No	Yes	No

Process influent mean TKN concentrations were in the range 19-22 mg N/L throughout Phases 1-5; no significant differences were observed between the A and B Side influent TKN, except during Phase 4, when the mean TKN concentration in the B side process influent (22.27 mg N/L) was significantly higher than that on the A Side

(20.81 mg N/L). Mean process TKN removal was correspondingly significantly higher on the B Side (20.29 mg N/L) than on the A Side (18.43 mg N/L) during Phase 4.

Mean TKN concentrations in unfiltered samples of the plant effluent were in the range 2-5 mg N/L, and did not differ significantly between the two sides, except during Phase 1, when the A Side mean (2.47 mg N/L) was significantly higher than the B Side mean (1.98 mg N/L). The difference was probably due to the significantly higher mean concentration of TSS in the plant effluent on the A Side during Phase 1, as shown earlier (see Table 6). Note that the mean TKN concentration in samples of filtered effluent did not differ between the A and B Sides during any of Phases 1-5, indicating that the experimental treatments carried out during Phases 1-5 had no significant effect on process TKN removal.

5.1.4 Ammonia

The results of the *t* test comparisons of the ammonia data for the A and B Side process trains during Phases 1-5 are summarized in Table 10. Similar to the BOD data described earlier (Table 8), reactor ammonia removals were normalized to show mg of ammonia removed per L of process influent (see Appendix 1). Mean process influent ammonia concentrations were in the range 11-15 mg N/L, and did not differ significantly between the two sides during Phases 1-5. Process effluent ammonia concentration for both the A and B Sides (not shown in Table 10) was consistently less than the detection limit of 0.01 mg N/L throughout the study. Process ammonia removal was not significantly different between the A and B Sides throughout Phases 1-5 (Table 10).

Table 10 - Results of Pilot Plant Monitoring - Ammonia

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Mean Process Influent	A Side	11.02	14.52	12.50	11.64	12.66
Ammonia Conc. (mg N/L)	B Side	10.98	13.99	12.55	11.56	12.87
Degrees of Freedom		10	12	24	20	9
Calculated t		0.52	1.80	-0.27	0.45	-0.39
Critical t (alpha=0.05)		2.23	2.18	2.06	20.9	2.26
Significant Difference?		No	No	No	No	No
Mean Ammonia Removal	A Side	11.02	14.52	12.70	11.03	12.86
Total Process (mg N/L)	B Side	10.84	13.98	12.76	12.16	12.66
Degrees of Freedom		10	12	22	21	9
Calculated t		.54	1.83	-0.26	-1.06	0.38
Critical t (alpha=0.05)		2.23	2.18	2.07	2.08	2.26
Significant Difference?		No	No	No	No	No
Mean Ammonia Removal	A Side	6.25	8.79	6.36	7.14	7.87
FGR 1 (mg N/L Influent)	B Side	5.95	7.52	5.46	5.81	6.91
Degrees of Freedom		9	12	22	21	9
Calculated t		0.47	4.25	4.12	1.92	1.71
Critical t (alpha=0.05)		2.26	2.18	2.07	2.08	2.26
Significant Difference?		No	Yes	Yes	No	No
Mean Ammonia Removal	A Side	2.78	1.85	3.61	3.24	4.81
FGR 2 (mg N/L Influent)	B Side	3.24	2.48	4.31	3.18	4.59
Degrees of Freedom		9	12	22	21	9
Calculated t		-1.30	-1.86	-5.06	0.14	0.62
Critical t (alpha=0.05)		2.26	2.18	2.07	2.08	2.26
Significant Difference?		No	No	Yes	No	No

The mean ammonia removal across each FGR cell is included in Table 10. As described in the literature review, the nitrification rate in trickling filters has been found be first order with respect to ammonia concentration where the ammonia concentration is 3-5 mg N/L or less, and zero order with respect to ammonia (oxygen limited) at higher ammonia concentrations. As shown earlier in Figure 1, each FGR was irrigated with the process liquid from its (completely mixed) individual underflow catchbasin. The mean concentrations of ammonia in the underflow catchbasins were in the range 1.6-3.0 mg N/L for FGR Cell 1, and 0.1-0.8 mg N/L for FGR Cell 2, throughout Phases 1-5. Corresponding ammonia removals across FGR Cells 1 and 2 were in the range 5.5-8.9 mg N/L of process influent and 1.2-3.0 mg

N/L of process influent, respectively (Table 10). Since the mean ammonia concentration in the influent to FGR Cell 1 was generally at or below the so-called first order threshold, and the mean ammonia concentration in the influent to FGR Cell 2 was always well below the first order threshold, it is likely that ammonia removal by the fixed growth in the FGRs was limited by ammonia flux, and not by dissolved oxygen flux (i.e., the ammonia removal rate was first order with respect to ammonia concentration). The higher mean ammonia removals in FGR Cell 1 compared to FGR Cell 2 confirm that the ammonia removal rate was lower at lower bulk solution ammonia concentrations. Residual ammonia in the effluent from FGR Cell 2 was removed to non-detectable levels in the aeration basin; mean mass removals of ammonia in the aeration basin (not shown in Table 10) were in the range 0.1-1.7 mg N/L of process influent (i.e., 1%-14% of total process ammonia removal), indicating that most of the process ammonia removal occurred in the FGRs.

Significant differences in aerobic mass removals of ammonia between the A and B Sides were not expected during Phases 1-3, since the experimental treatments associated with those phases involved changes to the anaerobic and anoxic reactors only. However, note that the mean mass of ammonia removal in the A Side FGR Cell 1 was significantly greater than that of the B Side during Phase 3. The lower mean mass of ammonia removal in the B Side FGR Cell 1 during Phase 3 resulted in a higher mean concentration of ammonia in the influent to the B Side FGR Cell 2 (0.83 mg N/L on the B Side compared to 0.59 mg N/L on the A Side), with a correspondingly greater mean mass of ammonia removal in FGR Cell 2 on the B Side than the A Side.

The FGR irrigation rates were always identical between the A and B Sides (24 L/min), except for during Phase 4, when the A Side FGRs received an irrigation rate of 24 L/min, compared to 14 L/min on the B Side. As shown in Table 10, the mean

ammonia removal in FGR Cell 1 during Phase 4 was 7.14 mg N/L on the A Side, compared to 5.81 mg N/L on the B Side; however, the difference was not statistically significant according to the t test, indicating the higher FGR irrigation rate did not result in a significantly greater degree of ammonia removal.

During Phase 5, when the A Side was operated at a significantly higher MLSS concentration than on the B Side, no significant difference in FGR mass ammonia removals was detected between the two sides (Table 10).

In summary, mean process influent ammonia concentrations in the range 11-15 mg N/L were consistently removed to non-detectable levels in both the A and B Side processes, with greater than 85% of the process ammonia removal occurring in the FGRs. Mean ammonia concentrations in the process mixed liquor cascaded over the FGRs were at or below the first order threshold, and ammonia removal in the FGRs was lower at lower bulk solution ammonia concentrations. An FGR irrigation rate of 14 L/min resulted in a lower mean mass of ammonia removal in FGR 1 than an irrigation rate of 24 L/min, but the difference was not statistically significant.

5.1.5 Nitrate and Nitrite

As described in the literature review, recycling of nitrates to the anaerobic reactor in biological phosphorus removal systems should be avoided, and UCT-type systems can be operated to achieve a near zero discharge of nitrite+nitrate (NO_x) to the anaerobic reactor. The t test comparisons of the mean concentration of NO_x in the return settled biosolids stream, the mean concentration of NO_x in the anoxic reactor, and the mean anoxic mass reduction of NO_x are summarized in Table 11. All reactor mass removals of NO_x were based on nominal retention time (see Appendix 1).

Table 11 - Monitoring of Pilot Plant Performance - Nitrate+Nitrite

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Mean NOx Conc. in Return	A Side	8.29	7.83	7.39	7.82	8.67
Settled Biosolids (mg N/L)	B Side	7.87	7.36	7.31	7.09	7.69
Degrees of Freedom		11	13	21	20	9
Calculated t		2.38	2.15	0.56	0.93	1.30
Critical t (alpha=0.05)		2.20	2.16	2.08	2.09	2.26
Significant Difference?		Yes	No	No	No	No
Mean Anoxic Reactor	A Side	0.21	0.64	0.51	0.32	0.23
NOx Conc. (mg N/L)	B Side	0.40	0.40	0.62	0.28	0.45
Degrees of Freedom		11	13	26	20	9
Calculated t		-2.81	1.82	-0.97	0.76	-2.29
Critical t (alpha=0.05)		2.20	2.16	2.06	2.09	2.26
Significant Difference?		Yes	No	No	No	Yes
Mean NOx Reduction	A Side	8.12	5.67	6.24	7.64	8.69
Anoxic (mg N/L Influent)	B Side	6.81	6.22	5.36	7.46	8.07
Degrees of Freedom		10	13	27	16	7
Calculated t		5.67	-1.54	1.67	0.24	1.06
Critical t (alpha=0.05)		2.23	2.16	2.05	2.12	2.36
Significant Difference?		Yes	No	No	No	No
Mean Specific NOx Reduction	A Side	0.97	0.80	0.99	1.06	1.32
Rate (mg N/hr/g anoxic MLSS)	B Side	1.28	0.70	0.98	0.91	1.77
Degrees of Freedom		10	13	20	17	7
Calculated t		-2.46	1.03	0.23	1.30	-2.59
Critical t (alpha=0.05)		2.23	2.16	2.09	2.12	2.36
Significant Difference?		Yes	No	No	No	Yes
Mean NOx Production	A Side	11.48	12.02	8.91	11.97	12.21
FGR 1 (mg N/L Influent)	B Side	9.57	10.89	7.28	10.19	10.84
Degrees of Freedom		10	13	27	16	7
Calculated t		2.68	2.68	4.52	3.52	1.41
Critical t (alpha=0.05)		2.23	2.16	2.05	2.12	2.36
Significant Difference?		Yes	Yes	No	Yes	No
Mean NOx Production	A Side	4.52	2.26	4.38	4.39	6.27
FGR 2 (mg N/L Influent)	B Side	5.13	2.86	5.04	5.80	6.11
Degrees of Freedom		10	13	27	16	7
Calculated t		-1.01	-1.42	-2.67	-2.53	0.16
Critical t (alpha=0.05)		2.23	2.16	2.05	2.12	2.36
Significant Difference?		No	No	Yes	Yes	No
Mean Process NOx Production	A Side	8.12	7.45	7.20	8.11	8.91
Incl. Denitrification (mg N/L)	B Side	7.68	7.14	6.83	7.95	8.51
Degrees of Freedom		10	13	27	16	7
Calculated t		1.83	1.18	1.60	1.00	6.04
Critical t (alpha=0.05)		2.23	2.16	2.05	2.12	2.36
Significant Difference?		No	No	No	No	Yes

During Phase 1, when the anoxic actual HRT on the A Side was 65 minutes compared to 35 minutes on the B Side, the mean NO_x concentration of 8.29 mg N/L in the return settled biosolids stream on the A Side was significantly higher than the mean of 7.87 mg N/L on the B Side, and the mean reduction of NO_x in the anoxic zone was correspondingly significantly higher on the A Side (8.12 mg N/L of process influent) than on the B Side (6.81 mg N/L of process influent - see Table 11). Despite the higher NO_x loading on the A Side, the mean NO_x concentration in the anoxic reactor was significantly lower on the A Side (0.21 mg N/L) than on the B Side (0.40 mg/N/L) during Phase 1, indicating that the lower anoxic HRT on the B Side resulted in a lower (but still adequate) degree of denitrification compared to the A Side.

Throughout Phases 2-4, the anoxic reactors on the A and B Sides both had an actual HRT of 35 minutes. No significant differences in the mean NO_x concentration in the return settled biosolids stream or in the anoxic reactor mass NO_x reduction were detected during Phases 2-4 (Table 11).

During Phase 5 (A Side anoxic mean MLSS concentration=3056 mg/L, compared to a mean of 2104 mg/L on the B Side), the A Side anoxic reactor mean NO_x concentration (0.23 mg N/L) was significantly lower than that of the B Side (0.45 mg N/L), despite a higher mean NO_x concentration in the return settled biosolids on the A Side (8.67 mg N/L) compared to the B Side (7.69 mg N/L). The mean mass reduction of NO_x in the anoxic reactor was correspondingly higher on the A Side (8.69 mg N/L of process influent) compared to the B Side (8.07 mg N/L of process influent) during Phase 5, although the difference was not statistically significant (Table 11).

Note that the anoxic mean MLSS concentration was significantly different between the A and B Sides during Phases 3 and 4, despite efforts to operate both sides at the same MLSS concentration (see Table 6). To factor out the effects of differences in anoxic operating MLSS concentration on mass removals of NO_x, the specific mass

removal rate of NO_x in the anoxic reactor was calculated, by dividing the anoxic reactor MLSS concentration into the anoxic mass reduction rate of NO_x, (anoxic reactor influent NO_x concentration minus reactor effluent NO_x concentration divided by reactor actual HRT). Note that, in cases where substrate removal was virtually complete within a time significantly less than the available HRT, the calculated specific rates of NO_x reduction would be less than the actual biochemical reaction rate (the batch test results discussed later in this thesis show that this was normally the case for NO_x removal in the anoxic reactor).

As shown in Table 11, the specific NO_x reduction rate during Phase 1 was significantly greater on the B Side than on the A Side; however, the lower calculated rate on the A Side can likely be attributed to the longer HRT on that side, and not to any difference in the biochemical reaction rates. The specific NO_x reduction rate was not significantly different between the two sides during Phases 2-4, when the anoxic HRT was equal on both sides. However, the specific NO_x reduction rate during Phase 5 was significantly higher on the B Side (1.77 mg N/hr/g MLSS) than on the A Side (1.32 mg N/hr/g MLSS), despite no significant difference in the reactor mass reduction between the two sides. Apparently, the anoxic HRT was sufficient for the lower mass of solids on the B Side to achieve a similar total mass of NOx reduction to the A Side. Comparisons of anoxic NO_x reduction rates are further discussed later in the section on batch test results.

The t test comparisons between the reactor mass balances for NO_x production in the FGRs are included in Table 11. The mean NO_x production in FGR Cells 1 and 2 was in the range 7-12 mg N/L of process influent, and 2-6 mg N/L of process influent, respectively. The mean production of NO_x across FGR Cells 1 and 2 followed a similar pattern to that observed for ammonia removal shown earlier in Table 10. That is, the NO_x production in FGR Cell 1 (mean influent ammonia concentration 1.3-3.0

mg N/L) was higher than that in FGR Cell 2 (mean influent ammonia concentration 0.1-0.8 mg N/L). Note that the mean production of NO_x across FGR Cells 1 and 2 (Table 11) was always greater than the corresponding mean removal of ammonia (Table 10), throughout Phases 1-5. The higher NO_x production was probably due to oxidation of ammonia which was produced by ammonification of organically-bound nitrogen in the process influent (note that the mean process TKN removals shown in Table 10 were in the range 18-22 mg N/L, compared to mean process influent ammonia concentrations in the range 10-15 mg N/L - see Table 8). The mean production of NO_x in the solids contact basin was in the range 0.00-1.34 mg N/L of process influent, confirming that most of the nitrification occurred in the FGRs.

The t test comparisons for the mean production of NO_x across the entire process (including the effects of denitrification), are included in Table 11. As shown, mean process NO_x production was in the range 7-9 mg N/L. No significant difference in NO_x production between the two sides was detected during Phases 1-4. During Phase 5, the mean NO_x production of 8.91 mg N/L on the A Side was significantly higher than the mean of 8.51 mg N/L on the B Side (the reason for the t test detecting a significant difference between two apparently similar means in this case was an extremely low variance in the data). As shown earlier in Table 10, mean process influent TKN concentration during Phase 5 was higher on the A Side (23.20 mg N/L) than on the B Side (22.33 mg N/L), although the difference was not statistically significant. Therefore, the slightly greater NO_x production on the A Side during Phase 5 was probably due to ammonification and nitrification of the higher TKN loading to the A Side.

In summary, the NO_x data showed that a reduction in anoxic reactor actual HRT from 65 minutes to 35 minutes caused a significant but relatively slight increase in the mean steady-state anoxic reactor NO_x concentration (i.e., an increase of 0.19 mg N/L).

Similarly, during Phase 5, the lower mean operating MLSS concentration on the B Side resulted in a significant but relatively slight increase in the mean steady-state anoxic reactor NO_x concentration, compared to the A Side (i.e., an increase of 0.22 mg N/L). Apparently, the anoxic actual HRT of 35 minutes was adequate for effective denitrification under all of the experimental conditions studied during Phases 1-5. The relationship between operating MLSS concentration and denitrification rate was investigated during the Phase 5 batch tests, discussed later in this thesis.

5.1.6 Volatile Fatty Acids

As described in the literature review, effective biological phosphorus removal depends on an adequate supply of short chain volatile fatty acids to the anaerobic reactor. As described earlier in Section 5.1.2., measurement of BOD_5 can provide some insight into the uptake of soluble carbonaceous substrates by phosphorus accumulating bacteria in the anaerobic phase; however, VFA concentrations provide a more accurate and relevant measurement.

The t test comparisons between the mean concentration of volatile fatty acids (VFA) in the process influent on the A and B Sides throughout Phases 1-5 are shown in Table 12. The t test comparisons between the mean VFA concentration in the anaerobic reactor, the mean mass of VFA uptake across the anaerobic reactor, and the mean specific VFA uptake across the anaerobic reactor, are also included in Table 12. The mass of VFA uptake is expressed in terms of mg HAc/L of process influent (i.e., uptake is based on nominal retention time, and not on actual retention time-see Appendix 1). The specific VFA uptake rate was calculated by dividing the MLSS concentration in the anaerobic reactor into the anaerobic mass VFA uptake rate (reactor influent concentration minus reactor effluent concentration divided by reactor actual retention time). The calculated specific rates of VFA uptake therefore do not necessarily give a true picture of the actual biochemical reaction rate; that is, in

cases where substrate removal had virtually ceased within a time significantly less than the available HRT, the calculated removal rate would be less than the actual biochemical reaction rate (the batch test results discussed later in this thesis show that this was often the case for VFA removal in the anaerobic reactor).

Table 12 - Monitoring of Pilot Plant Performance - Total Volatile Fatty Acids

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Process Influent Mean	A Side	26.84	39.84	36.75	34.71	44.17
Total VFA Conc. (mg/L)	B Side	21.70	39.21	35.64	34.61	41.59
Degrees of Freedom		11	13	21	20	9
Calculated t		3.37	0.65	0.85	0.10	1.53
Critical t (alpha=0.05)		2.20	2.16	2.08	2.09	2.26
Significant Difference?		Yes	No	No	No	No
Anaerobic Mean VFA	A Side	2.65	1.21	5.30	2.81	0.19
Conc. (mg/L Influent)	B Side	2.47	0.68	1.76	3.51	4.51
Degrees of Freedom		11	13	21	16	9
Calculated t		0.64	1.68	6.20	-0.50	-4.93
Critical t (alpha=0.05)		2.20	2.16	2.08	2.12	2.26
Significant Difference?		No	No	Yes	No	Yes
Anaerobic Mean VFA	A Side	24.80	37.67	23.35	27.03	43.60
Uptake (mg/L Influent)	B Side	18.22	42.34	32.19	23.97	27.96
Degrees of Freedom		10	13	21	16	9
Calculated t		2.69	-0.05	-3.64	0.76	3.75
Critical t (alpha=0.05)		2.23	2.16	2.08	2.12	2.26
Significant Difference?		Yes	No	Yes	No	Yes
Mean Specific Anaerobic VFA	A Side	21.84	22.08	60.79	45.70	47.99
Uptake Rate (mg/hr/g MLSS)	B Side	14.11	30.66	32.67	47.45	44.38
Degrees of Freedom		10	13	20	14	8
Calculated t		3.28	-3.41	5.40	-0.70	0.58
Critical t (alpha=0.05)		2.23	2.16	2.09	2.16	2.31
Significant Difference?		Yes	Yes	Yes	No	No

As shown in Table 12, the mean process influent VFA concentration was significantly higher on the A Side (26.84 mg HAc/L) than the B Side (21.70 mg HAc/L) during Phase 1. The reason for the difference is unknown. The mean concentration of VFA in both the A and B Side anaerobic reactors during Phase 1 was approximately 2.5 mg HAc/L. The mean anaerobic mass of VFA uptake on the A Side (24.80 mg HAc/L)

was significantly higher than on the B Side (18.22 mg HAc/L), indicating that the higher mass loading of VFA on the A Side resulted in a greater mass of VFA uptake compared to the B Side. The mean specific VFA uptake rate on the A Side (21.84 mg HAc/hr/g MLSS) was also significantly higher than on the B Side (14.11 mg HAc/hr/g MLSS) during Phase 1. Note that bulk solution VFA were typically not exhausted in either the A or B Side anaerobic reactors during Phase 1, indicating incomplete bacterial VFA uptake, despite an actual anaerobic retention time of 45 minutes on both sides (the VFA concentration in the anaerobic reactor was in the non-detectable range in only 4 out of the 12 samples collected for each of the A and B Sides). However, batch test results presented later in this thesis show that rapid VFA uptake typically ceased within the initial 25 minutes of the anaerobic phase.

During Phase 2, the process influent mean VFA concentration did not differ significantly between the two sides (39.84 mg HAc/L for the A Side and 39.21 mg HAc/L for the B Side - see Table 12). The mean anaerobic VFA uptake was lower on the A Side (37.67 mg HAc/L) than on the B Side (42.34 mg HAc/L) during Phase 2, despite a longer anaerobic actual HRT on the A Side (45 minutes), compared to the B Side (25 minutes), although the difference was not statistically significant. The mean VFA concentration in the A Side anaerobic reactor (1.21 mg HAc/L) was correspondingly greater than that on the B Side (0.68 mg HAc/L), but again the difference was not statistically significant (Table 12). Note that the mean specific VFA uptake rate on the A Side (22.08 mg HAc/hr/g MLSS) was significantly lower than that on the B Side (30.66 mg HAc/hr/g MLSS) on the B Side during Phase 2. The lower calculated specific VFA uptake rate on the A Side can be attributed to the longer anaerobic HRT on that side during Phase 2 (see Table 6). In any case, since the B Side had the shorter anaerobic HRT during Phase 2, the data imply that reducing the anaerobic HRT from 45 minutes to 25 minutes had no detrimental effect on the mass of carbon stored in the anaerobic phase. Anaerobic bulk solution VFA were

removed to non-detectable levels in 9 out of 14 samples on the A Side, and in 7 out of 14 samples on the B Side during Phase 2.

As shown in Table 12, the process influent mean VFA concentration was not significantly different between the two sides during Phase 3 (36.75 mg HAc/L on the A Side and 35.64 mg HAc/L on the B Side). The anaerobic mean VFA uptake of 32.19 mg HAc/L on the B Side (25 minute actual HRT) was significantly greater than the mean uptake of 23.35 mg HAc/L on the A Side (8 minute actual HRT) during Phase 3, indicating that the shorter anaerobic HRT on the A Side resulted in a significant decrease in anaerobic carbon storage, compared to the B Side (Table 12). The mean VFA concentration in the anaerobic reactor was correspondingly significantly higher on the A Side (5.30 mg HAc/L) than on the B Side (1.76 mg HAc/L). The anaerobic mean specific VFA uptake rate of 60.79 mg HAc/hr/g MLSS on the A Side was significantly higher than the mean specific VFA uptake rate of 32.67 mg HAc/g MLSS on the B Side. However, similar to the results discussed above for Phase 2, batch test results presented later in this thesis show that rapid VFA uptake on the B Side typically ceased within less than the available 25 minute HRT; therefore, the calculated VFA uptake rate on the B Side is likely lower than the actual biochemical reaction rate. The bulk solution VFA concentration was reduced to non-detectable levels in none of the 22 samples on the A Side, and in 7 out of 22 samples on the B Side during Phase 3.

As shown in Table 12, there was no significant difference between the process influent VFA concentration, the mass of VFA uptake, the specific VFA uptake rate, or the VFA concentration in the anaerobic reactor between the A and B Sides during Phase 4, when both sides were operated with identical anaerobic HRTs (i.e., 8 minutes for Phase 4a, and 25 minutes for Phase 4b). However, note that the mean anaerobic reactor MLSS concentration was significantly lower on the A Side (2215

mg/L) than on the B Side (2590 mg/L) during Phase 4. Despite the lower operating mean MLSS concentration on the A Side, the mean mass of VFA uptake on the A Side (27.03 mg HAc/L) was higher than that of the B Side (23.97 mg HAc/L). The mean concentration of VFA in the anaerobic reactor was correspondingly lower on the A Side (2.81 mg HAc/L) than on the B Side (3.51 mg HAc/L). The mean specific anaerobic VFA uptake rate on the A Side (45.70 mg HAc/hr/g MLSS) was lower than that of the B Side (47.45 mg HAc/hr/g MLSS-see Table 12). During Phase 4a (8 minute anaerobic HRT), bulk solution VFA were reduced to non-detectable levels in none of 9 samples on the A Side, and in 1 out of 9 samples on the B Side. During Phase 4b (25 minute anaerobic HRT), bulk solution VFA were reduced to non-detectable levels in 2 out of 8 samples on the A Side, and in 8 out of 8 samples on the B Side.

During Phase 5, both sides were operated with an anaerobic actual HRT of 25 minutes. As described earlier (Chapter 4), the B Side was operated at a significantly lower anaerobic mean MLSS concentration (1464 mg/L) than the A Side (2121 mg/L) during Phase 5. The mean process influent VFA concentration did not differ significantly between the two sides during Phase 5 (Table 12). However, the A Side anaerobic mean VFA uptake (43.60 mg HAc/L) was significantly higher than that of the B Side (27.96 mg HAc/L). The A Side anaerobic reactor mean VFA concentration (0.19 mg HAc/L) was correspondingly significantly lower than that of the B Side (4.51 mg HAc/L). The mean specific VFA uptake rate did not differ significantly between the two sides, although the A Side (47.99 mg HAc/hr/g MLSS) was slightly higher than the B Side (44.38 mg HAc/hr/g MLSS). Bulk solution VFA were reduced to non-detectable levels in 9 out of 9 samples on the A Side, and in only 2 out of 9 samples on the B Side during Phase 5.

In summary, the reduction in anaerobic actual HRT from 45 minutes to 25 minutes

(Phase 2) had no significant effect on the mean mass of VFA taken out of solution in the anaerobic reactor. However, a further reduction in anaerobic actual HRT to 8 minutes (Phase 3) caused a significant decrease in anaerobic mean VFA uptake, with a corresponding increase in anaerobic reactor steady-state mean VFA concentration. During the investigation of the effects of operating MLSS concentration (Phase 5), an anaerobic mean MLSS concentration of 1,464 mg/L resulted in a significantly lower mean mass of anaerobic VFA uptake than an anaerobic mean MLSS concentration of 2,121 mg/L; however, the calculated specific VFA uptake rate was not significantly different between the two sides, indicating that the lower mass of VFA uptake at the lower operating MLSS concentration was simply due to a lower mass of solids in the anaerobic reactor. Where the anaerobic actual HRT was 25 minutes or more, bulk solution VFA in the anaerobic reactor were reduced to non-detectable levels in 24 out of 43 samples on the A Side, and in 28 out of 65 samples on the B Side. Where the anaerobic actual HRT was 8 minutes, bulk solution anaerobic VFA concentrations were typically greater than zero. The implications of significant steady-state VFA concentrations in the anaerobic reactor are discussed later in this thesis.

5.1.7 Phosphorus

The t test comparisons between the A and B Side process influent mean total phosphorus concentration, process mass removal of total phosphorus, plant effluent orthophosphate (PO_4^{3-}) concentration, and the percent phosphorus by dry weight in the aeration basin MLSS are summarized in Table 13. As shown, process influent mean total phosphorus concentration was in the range 3.2-4.0 mg P/L, and did not differ significantly between the A and B Sides throughout Phases 1-5.

Table 13 - Monitoring of Pilot Plant Performance - Total Phosphorus

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Mean Process Influent Total	A Side	3.28	3.84	3.99	4.01	3.97
Phosphorus Conc. (mg P/L)	B Side	3.25	3.93	3.90	3.99	3.91
Degrees of Freedom		11	13	26	21	8
Calculated t		0.23	-0.32	0.74	0.02	0.52
Critical t (alpha=0.05)		2.20	2.16	2.06	2.08	2.31
Significant Difference?		No	No	No	No	No
Mean Process Removal of	A Side	1.97	3.09	3.65	2.00	3.60
Total Phosphorus[1] (mg P/L)	B Side	2.39	3.84	3.56	2.56	1.81
Degrees of Freedom		10	10	21	21	8
Calculated t		-1.92	-1.82	0.45	-1.93	7.45
Critical t (alpha=0.05)		2.23	2.23	2.08	2.08	2.31
Significant Difference?		No	No	No	No	Yes
Mean Plant Effluent	A Side	0.62	0.53	0.27	1.86	0.38
PO_4^{3-} Conc. (mg P/L)	B Side	0.38	0.17	0.10	1.26	1.73
Degrees of Freedom		11	13	22	20	9
Calculated t		1.88	3.57	1.83	2.91	-6.86
Critical t (alpha=0.05)		2.20	2.16	2.07	2.09	2.26
Significant Difference?		No	Yes	No	Yes	Yes
Mean % Phosphorus by Wt.	A Side	3.7	4.4	3.9	3.8	4.6
in Dried Aeration Basin MLSS	B Side	3.3	4.1	4.3	4.1	3.4
Degrees of Freedom		11	13	27	21	9
Calculated t		2.64	2.05	-2.89	-2.26	4.86
Critical t (alpha=0.05)		2.20	2.16	2.05	2.08	2.26
Significant Difference?		Yes	No	Yes	Yes	Yes

[1] based on filtered effluent samples

Mean removal of total phosphorus did not differ significantly between the two sides during Phases 1-3, indicating that the progressive reductions in the anoxic and anaerobic reactor HRTs did not have a significant effect on total phosphorus removal. Similarly, the lower FGR recycle rate on the B Side during Phase 4 did not result in a significantly lower mean removal of total phosphorus, compared to the A Side (Table 13). During Phase 5, process mean removal of total phosphorus was significantly higher on the A Side (3.60 mg P/L), compared to the B Side (1.81 mg P/L), indicating that the lower operating MLSS concentration on the B Side during Phase 5 resulted in a significantly lower degree of phosphorus removal, compared to the A Side. It

should be noted that, due to time constraints, there was no acclimation period between Phases 4 and 5, and the pilot plant was only operated for one month during Phase 5. After a longer period of acclimation, the percentage of phosphorus accumulating bacteria in the B Side process MLSS might have increased, and good phosphorus removal might have been achieved at the lower operating MLSS concentration.

The mean orthophosphate concentration in the plant effluent for both the A and B Sides was in the range 0.10-0.62 mg P/L, during Phases 1-3 (Table 13). The t test did not detect a significant difference in effluent mean orthophosphate concentration between the A Side (0.62 mg P/L) and the B Side (0.38 mg P/L) during Phase 1, when the anoxic actual HRTs on the A and B Sides were 65 minutes and 35 minutes, respectively. During Phase 2, the effluent mean orthophosphate concentration on the A Side (0.53 mg P/L) was significantly higher than that of the B Side (0.17 mg P/L), despite a longer anaerobic actual HRT on the A Side (45 minutes) than on the B Side (25 minutes). During Phase 3, when the anaerobic actual HRTs on the A and B Sides were 8 minutes and 25 minutes, respectively, the mean effluent orthophosphate concentration on the A Side (0.27 mg P/L) was not significantly different from that of the B Side (0.10 mg P/L). The mean effluent orthophosphate concentration was significantly higher on the A Side (1.86 mg P/L) than on the B Side (1.26 mg P/L) during Phase 4. Note that the mean effluent orthophosphate concentration for both sides during Phase 4 was much higher than during Phases 1-3. During Phase 5, the mean effluent orthophosphate concentration was significantly lower on the A Side (0.38 mg P/L) than on the B Side (1.73 mg P/L), confirming that the lower operating MLSS concentration on the B Side resulted in a significantly lower degree of phosphorus removal (Table 13). However, as discussed earlier, the Phase 5 results should be considered in light of the fact that there was no acclimation period between Phases 4 and 5.

The phosphorus content of the process suspended solids is an indicator of the presence of phosphorus storing bacteria in biological treatment systems. The suspended solids in conventional systems not designed for biological phosphorus removal generally contain only that phosphorus which is required for normal metabolism and growth, usually 1.5-2% phosphorus by dry weight. The t test comparisons of the phosphorus content of the aeration basin MLSS are included in Table 13. The mean phosphorus content of the aeration basin MLSS was in the range 3%-5%, indicating a significant degree of excess bacterial storage of phosphorus.

During Phase 1, the mean phosphorus content by dry weight of the aeration basin MLSS in the A Side process was 3.7%, significantly higher than the phosphorus content of 3.3 % on the B Side (note that the aeration basin MLSS concentration was significantly higher on the B Side during Phase 1 - see Table 6). The phosphorus content was also higher on the A Side (4.4%) than on the B Side (4.1%) during Phase 2, when the A Side anaerobic actual HRT was 45 minutes on the A Side, compared to 25 minutes on the B Side, although in this case the difference was not statistically significant. Phase 3 operation compared an anaerobic actual HRT of 8 minutes on the A Side to 25 minutes on the B Side; the A Side aeration basin MLSS contained significantly less phosphorus during this period (3.9%) than did the B Side (4.3%). The experimental difference between the two sides during Phase 4 was an FGR recycle rate of 24 L/min on the A Side, compared to 14 L/min on the B Side; during Phase 4, the B Side aeration basin MLSS phosphorus content (4.1%) was significantly higher than that of the A Side (3.8%). For Phase 5, the aeration basin MLSS on the A Side (mean aeration basin MLSS concentration = 3089 mg/L) contained 4.6% phosphorus by dry weight, significantly more than the mean phosphorus content of 3.4% on the B Side (mean aeration basin MLSS concentration = 1965 mg/L). Note a higher MLSS phosphorus content (Table 13) is generally associated with a higher anaerobic specific VFA mass uptake (see Table 12 in the previous section). The

implications of the phosphorus content of the process MLSS on VFA uptake and phosphorus removal will be discussed later in this thesis.

Process influent mean total phosphorus concentration and process effluent orthophosphate concentration for the A and B Side process trains throughout Phases 1-5 are summarized in Figures 3a and 3b, respectively. Some of the short-term increases in effluent orthophosphate shown in Figure 3 can be attributed to equipment failures. However, following a period of effective phosphorus removal from January through July of 1993 (which included Phases 1 and 2), plant effluent orthophosphate concentration rose from values typically less than 1 mg P/L to greater than 2 mg P/L (see Figure 3). The increase in orthophosphate was not associated with equipment failures, and effluent parameters other than phosphorus did not increase.

As described earlier, investigations during November of 1993 revealed an excessive accumulation of solids on the FGR media, caused by suspended organisms in the mixed liquor adhering to the biofilm on the media. The hydraulic pulse loading regime introduced in late November of 1993 on both the A and B Sides to flush accumulated solids from the media resulted in an immediate and consistent improvement in biological phosphorus removal in both process trains (Figure 3). Note that phosphorus removal on the A Side process was more erratic than on the B Side during Phases 1 and 2; as described earlier, solids accumulation with subsequent uncontrolled sloughing events during this period was more severe on the A Side, and this might explain the significantly higher plant effluent mean orthophosphate concentration on the A Side compared to the B Side during Phase 2 (Table 13).

Following the completion of Phase 3 in late February of 1994, the actual HRT of the B Side anaerobic reactor was reduced from 25 minutes to 8 minutes, to match that of the A Side. Note that during Phase 3, phosphorus removal on the A Side was effective with an 8 minute anaerobic HRT for a period of approximately 3 months, from late

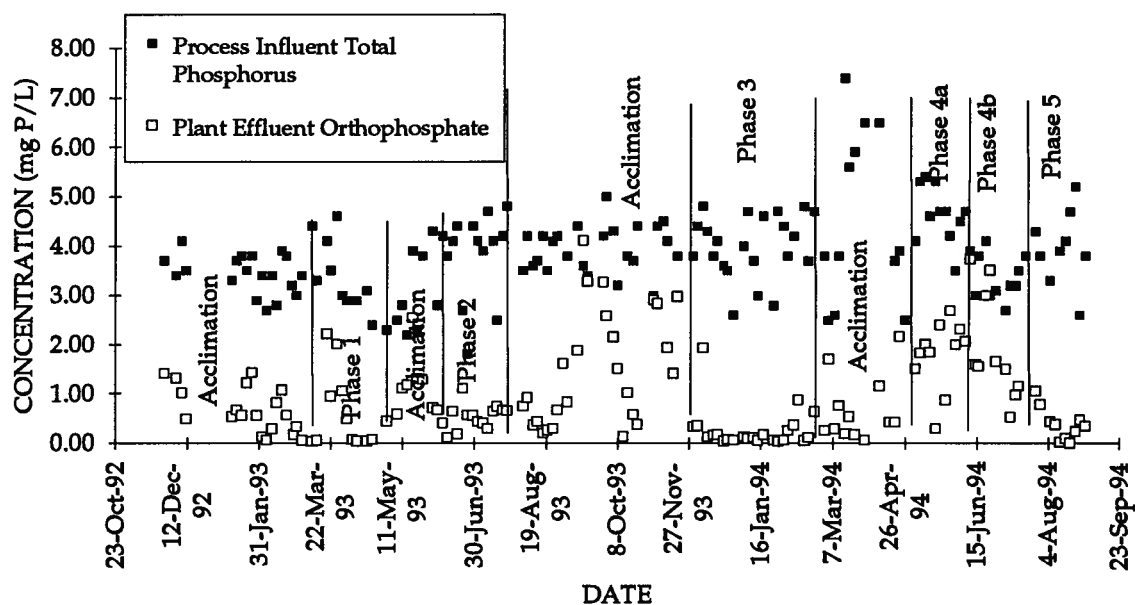November of 1993 through late February of 1994 (Figure 3a).

Figure 3a - Pilot Plant Phosphorus Removal - A Side Process Train
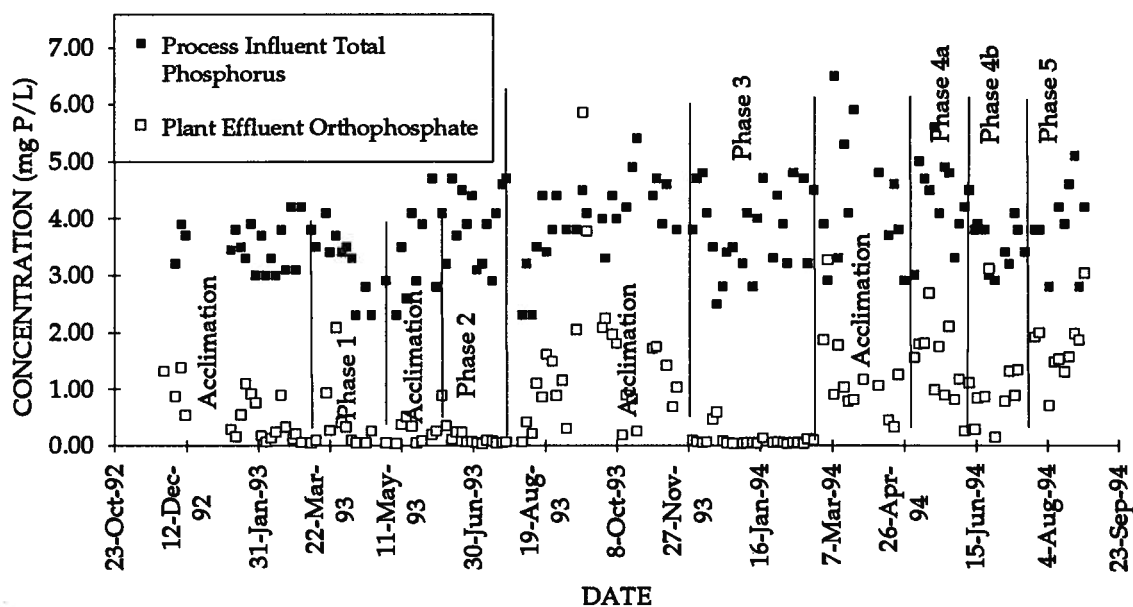
Figure 3b - Pilot Plant Phosphorus Removal - B Side Process Train

However, following the reduction in anaerobic HRT on the B Side, an immediate deterioration in phosphorus removal was observed (Figure 3b). At approximately the same time, phosphorus removal also began to deteriorate on the A Side (Figure 3a). Therefore, it appeared that a change had occurred near the end of Phase 3, which rendered the 8 minute anaerobic HRT insufficient for good phosphorus removal. In mid-June of 1994, the anaerobic actual HRT was increased to 25 minutes on both the A and B Sides, and effluent orthophosphate concentrations on both sides began to decrease (Figures 3a and 3b). On the B Side, the sludge wasting rate was increased near the end of July, 1994, to reduce the aeration basin operating MLSS concentration to approximately 2,000 mg/L, and effluent orthophosphate concentrations began to increase again (Figure 3b). Reasons for the deteriorations and improvements in phosphorus removal following phase 3 are discussed later in this thesis.

The *t* test comparisons between the A and B Side reactor mass balances for orthophosphate uptake and/or release throughout Phases 1-5 are summarized in Table 14. The mean uptake or release of orthophosphate was normalized to give mg P/L of process influent (i.e., the mass removals were based on nominal retention time, and not on actual retention time - see Appendix 1). Where the specific release or uptake rate of orthophosphate is shown, the value was calculated by dividing the reactor MLSS concentration into the mass release or uptake rate (reactor influent concentration minus reactor effluent concentration divided by reactor actual HRT). As described earlier, in cases where substrate release or uptake had ceased or slowed significantly within a time significantly less than the available HRT, the calculated release or uptake rate would be lower than the actual biochemical reaction rate (the batch test results discussed later in this thesis show that this was often the case for orthophosphate release in the anaerobic reactor and orthophosphate uptake in the anoxic and aerobic reactors).

Table 14 - Pilot Plant Performance - Reactor Orthophosphate Mass Balances

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Anaerobic Mean PO_4^{3-}	A Side	24.96	39.81	16.90	23.42	36.96
Release (mg P/L Influent)	B Side	23.68	41.16	32.97	21.63	21.64
Degrees of Freedom		10	13	21	21	9
Calculated t		0.86	-0.76	-15.63	0.58	7.78
Critical t (alpha=0.05)		2.23	2.16	2.08	2.08	2.26
Significant Difference?		No	Yes	Yes	No	Yes
Anaerobic Mean Specific PO_4^{3-}	A Side	7.38	8.41	15.06	13.13	13.74
Rel. Rate(mg P/hr/g MLSS)	B Side	6.13	11.36	10.81	15.81	11.42
Degrees of Freedom		10	13	20	19	8
Calculated t		3.68	-3.23	3.93	-1.87	2.74
Critical t (alpha=0.05)		2.23	2.16	2.09	2.09	2.31
Significant Difference?		Yes	Yes	Yes	No	Yes
Anoxic Mean PO_4^{3-}	A Side	13.00	19.56	3.56	10.54	8.29
Uptake (mg P/L Influent)	B Side	11.14	24.89	14.19	7.85	7.35
Degrees of Freedom		10	13	21	21	9
Calculated t		2.58	-4.85	-10.59	0.92	1.15
Critical t (alpha=0.05)		2.23	2.16	2.08	2.08	2.26
Significant Difference?		Yes	Yes	Yes	No	No
Anoxic Mean Specific PO_4^{3-}	A Side	1.49	2.70	0.47	1.59	1.31
Up. Rate (mg P/hr/g MLSS)	B Side	1.94	2.82	2.06	0.99	1.57
Degrees of Freedom		11	13	21	21	9
Calculated t		-2.67	-0.39	-10.23	1.39	-1.55
Critical t (alpha=0.05)		2.20	2.16	2.08	2.08	2.26
Significant Difference?		Yes	No	Yes	No	No
FGR Cell 1 Mean PO_4^{3-}	A Side	4.58	8.37	7.68	5.13	14.78
Uptake (mg P/L Influent)	B Side	4.90	9.90	9.50	5.60	4.33
Degrees of Freedom		10	13	21	21	9
Calculated t		-0.74	-1.68	-3.84	-0.54	4.37
Critical t (alpha=0.05)		2.23	2.16	2.08	2.08	2.26
Significant Difference?		No	No	Yes	No	Yes
FGR Cell 2 Mean PO_4^{3-}	A Side	2.22	3.83	2.87	1.96	8.78
Uptake (mg P/L Influent)	B Side	2.65	2.32	5.07	3.13	4.65
Degrees of Freedom		10	13	21	21	9
Calculated t		-1.29	0.76	-9.07	-2.33	2.35
Critical t (alpha=0.05)		2.23	2.16	2.08	2.08	2.26
Significant Difference?		No	No	Yes	No	Yes
Aeration Basin Mean PO_4^{3-}	A Side	6.56	10.09	5.47	6.17	8.78
Uptake (mg P/L Influent)	B Side	6.83	6.74	6.99	6.18	4.65
Degrees of Freedom		10	13	21	21	9
Calculated t		-0.33	1.93	-1.98	-0.02	2.35
Critical t (alpha=0.05)		2.23	2.16	2.08	2.08	2.26
Significant Difference?		No	No	No	No	Yes

Table 14 (cont.) - Pilot Plant Performance - Reactor Orthophosphate Mass Balances

Parameter		Phase 1	Phase 2	Phase 3	Phase 4	Phase 5
Aeration Mean Specific PO_4^{3-}	A Side	0.81	0.98	0.50	0.52	0.61
Up. Rate (mg P/hr/g MLSS)	B Side	0.66	0.45	0.70	0.47	0.73
Degrees of Freedom		11	13	21	21	9
Calculated t		2.12	2.75	-2.26	0.95	-2.21
Critical t (alpha=0.05)		2.20	2.16	2.08	2.08	2.26
Significant Difference?		No	Yes	Yes	No	No
Final Clarifier Mean PO_4^{3-}	A Side	-0.12	-0.27	0.20	-0.57	-0.52
Release (mg P/L Influent)	B Side	-0.17	0.00	0.17	-0.27	-0.74
Degrees of Freedom		10	13	21	21	9
Calculated t		0.34	-1.34	0.11	-1.02	1.66
Critical t (alpha=0.05)		2.23	2.16	2.08	2.08	2.26
Significant Difference?		No	No	No	No	No

In the anaerobic reactor, the mean mass release of orthophosphate did not differ significantly between the two sides during Phase 1 (24.96 mg P/L on the A Side and 23.68 mg P/L on the B Side), when the anaerobic actual HRT was 45 minutes for both sides (Table 14); however, the specific orthophosphate release rate was significantly higher on the A Side (7.38 mg P/hr/g MLSS) than on the B Side (6.13 mg P/hr/g MLSS). Note that the mean VFA loading to the anaerobic reactor was significantly higher on the A Side during Phase 1 (Table 12).

During Phase 2, the mean anaerobic orthophosphate release on the A Side (39.81 mg P/L) was not significantly different from that of the B Side (41.16 mg P/L), despite an anaerobic actual HRT of only 25 minutes on the B Side, compared to 45 minutes on the A Side. The calculated mean anaerobic specific orthophosphate release rate was significantly higher on the B Side (11.36 mg P/hr/g MLSS) than on the A Side (8.41 mg P/hr/g MLSS) during Phase 2; however, the difference was likely due to the higher HRT on the A Side, as described for specific VFA uptake in the previous section.

During Phase 3, when the A Side anaerobic actual HRT was 8 minutes compared to 25 minutes on the B Side, mean anaerobic orthophosphate mass release was significantly lower on the A Side (16.90 mg P/L) than on the B Side (32.97 mg P/L). The specific orthophosphate release rate of 15.06 mg P/hr/g MLSS on the A Side was significantly higher than that on the B Side (10.81 mg P/hr/g MLSS); again, the difference was likely due to the shorter HRT on the A Side, as described in the previous section on specific VFA release. During Phase 4, when the A and B Side anaerobic HRTs were equal, there was no significant difference in mean anaerobic orthophosphate mass release or specific release rate between the two sides (Table 14).

During Phase 5, anaerobic orthophosphate mass release and specific release rate were significantly higher on the A Side (36.96 mg P/L and 13.74 mg P/hr/g MLSS, respectively) compared to the B Side (21.64 mg P/L and 11.42 mg P/hr/g MLSS, respectively). Note that the mean anaerobic reactor operating MLSS concentration was 2121 mg/L on the A Side compared to only 1464 mg/L on the B Side, during Phase 5, and the lower specific orthophosphate release rate was associated with the lower mean MLSS concentration.

During Phase 1, when the A Side anoxic actual HRT was 65 minutes compared to only 35 minutes on the B Side, the mean uptake of orthophosphate in the anoxic reactor was significantly higher on the A Side (Table 14). As described later in this thesis, phosphorus uptake generally occurred during denitrification in the anoxic reactor, and the greater anoxic phosphorus removal on the A Side during Phase 1 can be attributed to the higher NO_x loading on the A Side at that time (see Table 11). The calculated mean specific anoxic orthophosphate uptake rate was significantly lower on the A Side (1.49 mg P/hr/g MLSS) than on the B Side (1.94 mg P/hr/g MLSS) during Phase 1; this can likely be attributed to the longer HRT on the A Side for Phase 1, in a similar manner to that described earlier for specific orthophosphate release

during Phases 2 and 3.

During Phases 2 and 3, the anoxic mean uptake of orthophosphate was significantly higher on the B Side, despite an equal anoxic HRT (35 minutes) on both sides. However, the mean specific anoxic orthophosphate uptake rate was not significantly different between the two sides during Phase 2. During Phase 3, the mean specific anoxic orthophosphate uptake was significantly higher on the B Side (2.06 mg P/hr/g MLSS) than on the A Side (0.47 mg P/hr/g MLSS). The lower specific anoxic orthophosphate uptake on the A Side during Phase 3 might be attributed to anoxic phosphorus release induced by the relatively high VFA concentrations in the process liquid leaving the A Side anaerobic reactor, which had an actual HRT of only 8 minutes. As described later in this thesis, simultaneous denitrification and phosphorus release under anoxic conditions were frequently observed in batch tests when high concentrations of VFA were allowed to enter the anoxic phase.

During Phase 4, when both sides had an anoxic actual HRT of 35 minutes, mean anoxic orthophosphate mass uptake and specific uptake rate did not differ significantly between the two sides. Similarly, during Phase 5, when the anoxic actual HRT was 35 minutes on both sides and the operating mean MLSS concentration was 3056 mg/L on the A Side compared to 2104 mg/L on the B Side, there was no significant difference between the mean anoxic orthophosphate mass uptake or mean specific uptake rate.

During Phases 1, 2, and 4 , mean orthophosphate uptake across FGR Cells 1 and 2 did not differ significantly between the two sides (Table 14). During Phase 3, orthophosphate uptake across FGR Cells 1 and 2 was significantly higher on the B Side than on the A Side. During Phase 5, mean orthophosphate uptake across FGR Cells 1 and 2 was significantly higher on the A Side.

As shown in Table 14, there was no significant difference in the mean mass of orthophosphate uptake in the aeration basin during Phases 1-4. During Phase 5, the A Side uptake (8.78 mg P/L) was significantly higher than that of the B Side (4.65 mg P/L). The specific mass rate of orthophosphate uptake in the aeration basin is also compared in Table 14. As shown, the mean specific orthophosphate uptake rate in the aeration basin did not differ significantly between the two sides during Phase 1. During Phase 2, the mean specific orthophosphate uptake rate was significantly higher on the A Side (0.98 mg P/hr/g MLSS) than on the B Side (0.45 mg P/hr/g MLSS). During Phase 3, the B Side mean specific aerobic orthophosphate uptake rate on the B Side (0.70 mg P/hr/g MLSS) was significantly higher than that of the A Side (0.50 mg P/hr/g MLSS). The mean specific orthophosphate uptake rate across the aeration basin did not differ significantly between the two sides during Phases 4 and 5 (Table 14).

As shown in Table 14, a mean release of orthophosphate in the secondary clarifier was observed only during Phase 3 (0.20 mg P/L on the A Side and 0.17 mg P/L on the B Side). During Phases 1, 2, 4, and 5, a mean orthophosphate uptake in the range 0.1-0.8 mg P/L in the secondary clarifier was observed. The t test did not detect a significant difference between the A and B Side means for orthophosphate release or uptake in the secondary clarifier during any of Phases 1-5.

In summary, the reduction in anoxic actual HRT from 65 minutes to 35 minutes (Phase 1) did not result in a statistically significant change in plant phosphorus removal, although the mean process removal of total phosphorus was higher and plant effluent mean orthophosphate concentration was lower at the lower anoxic HRT (Table 13). Progressive reductions in the anaerobic actual HRT from 45 minutes to 25 minutes (Phase 2), and from 25 minutes to 8 minutes (Phase 3) did not cause significant changes in mean process total phosphorus removal, although the 8 minute

HRT resulted in a significantly lesser degree of anaerobic VFA uptake, compared to the 25 minute HRT. Note that plant effluent orthophosphate concentration was significantly lower at the 25 minute anaerobic HRT, compared to the 45 minute HRT (Table 13). As described earlier, following the completion of Phase 3, it became apparent that a change or changes in process parameters had occurred, which rendered the 8 minute anaerobic HRT inadequate for effective phosphorus removal; possible reasons for the deterioration in phosphorus removal following Phase 3 will be discussed later in this thesis. The comparison of FGR recycle rates during Phase 4 showed that phosphorus removal was not significantly different between the two sides. The investigation of the effects of operating MLSS concentration (Phase 5) showed that phosphorus removal was much more effective at an aeration basin mean operating MLSS concentration of 3089 mg/L than at a mean concentration of 1965 mg/L. Throughout Phases 1-5, significant quantities of phosphorus were removed from solution in the anoxic reactor (typically 7-25 mg P/L). Daily hydraulic pulse loading of the FGRs to prevent solids accumulation on the FGR media greatly improved process phosphorus removal.

5.1.8 Actual HRTs Through the FGR Media

The actual HRT of the process liquid in the FGR media was estimated by the method described in Chapter 4. The results of the testing to estimate the actual HRT of the process liquid on the FGR media are summarized in Table 15. The nominal HRTs shown in Table 15 were estimated by multiplying the actual HRT by the ratio of the FGR irrigation rate:process influent flow rate, where the process influent flow rate was assumed to be 4.8 L/min.

The results shown in Table 15 for November 8 and 9 of 1993 were obtained during the acclimation period between Phases 2 and 3 (before the institution of hydraulic pulse loading of the FGRs), when the steady-state irrigation rate for all four FGR cells was

24 L/min. The results shown for July 19, 1994 (after the institution of pulse loading), were obtained during Phase 4, when the steady-state FGR irrigation rate was 24 L/min for the A Side FGR Cells 1 and 2, and 14 L/min for the B Side Cells 1 and 2. In all cases, the calculated average actual HRT was in the range 0.8-1.2 minutes, and the nominal HRT was in the range 3.6-6.5 minutes (Table 15).

Table 15 - FGR Actual HRTs

FGR Cell	Date of Test	FGR Irrigation Rate (L/min)	Liquid Volume Collected (L)	Average Actual HRT (min)	Average Nominal HRT (min)
A Side Cell 1	Nov 9/93	26.1	26.5	1.0	5.4
	Nov 9/93	24.5	24.5	1.0	5.1
	Jul 19/94	24.2	18.7	0.8	4.0
B Side Cell 1	Nov 9/93	24.7	27.5	1.1	5.7
	Nov 9/93	24.8	27.0	1.1	5.7
	Jul 19/94	15.0	18.3	1.2	3.8
A Side Cell 2	Nov 8/93	24.0	31.0	1.3	6.5
	Nov 8/93	24.3	28.5	1.2	6.1
	Jul 19/94	24.5	24.9	1.0	5.1
	Jul 19/94	24.5	24.0	1.0	5.1
B Side Cell 2	Nov 8/93	26.1	30.5	1.2	6.5
	Nov 8/93	24.5	29.0	1.2	6.1
	Jul 19/93	14.2	17.5	1.2	3.6

According to the results shown in Table 15, the lower FGR irrigation rate on the B Side during Phase 4 (test date July 19, 1994) might have resulted in a slightly longer actual HRT, compared to the A Side. For Cell 1 on July 19, 1994, the A Side actual HRT was 0.8 minutes, compared to 1.2 minutes on the B Side. Similarly, for Cell 2 on July 19, the A Side actual HRT was 1.0 minutes, compared to 1.2 minutes on the B Side. However, the calculated nominal HRT during Phase 4 was longer on the A Side (4.0-5.1 minutes) than on the B Side (3.6-3.8 minutes), due to the higher FGR recycle rate on the A Side (5 times the process influent flow rate), compared to the B Side (3 times the process influent flow rate). The actual retention times before the institution of FGR pulse loading for the A Side (1.0-1.3 min) were slightly longer than those

recorded after pulse loading was begun (0.8-1.0 min), indicating that hydraulic scouring of the media reduced the actual HRT of the mixed liquor through the tower.

5.2 BATCH TEST RESULTS

Bench-scale batch tests designed to simulate the performance of the suspended growth process reactors were periodically conducted, to evaluate biochemical reaction rates. As described in Chapter 4, the batch tests simulated a plug flow regime, while the pilot plant reactors were completely mixed, flow through basins. The batch tests therefore simulated the pilot plant reactors only in cases where the substrate reaction rate was observed to be zero order with respect to substrate concentration.

As described in the previous section, pilot plant performance was effective at mixed liquor temperatures as low as 9 °C. However, the batch tests were all conducted at room temperature (approximately 20 °C); the experimental design did not include an investigation of the effects of temperature on substrate removal rates.

5.2.1 Phase 1 - 65 Minute Anoxic Actual HRT vs. 35 Minute Anoxic Actual HRT

As shown in Table 3, Phase 1 operation compared an anoxic actual HRT of 65 minutes on the A Side to 35 minutes on the B Side. Two batch tests were conducted during Phase 1. The results of one of the Phase 1 batch tests (conducted on April 29, 1993) are shown in Figures 4a (A Side simulation) and 4b (B Side simulation).

In the anoxic phase on the A Side, denitrification of nitrate was accompanied by phosphorus uptake. Denitrification and phosphorus uptake were completed within the first 30 minutes, with orthophosphate release being observed over the remaining 35 minutes of the test period (Figure 4a). The results imply that the bacteria responsible for enhanced biological phosphorus removal were capable of using

nitrate rather than oxygen as an electron acceptor for utilization of stored carbon and associated phosphorus uptake-storage. On the B Side, denitrification with associated phosphorus uptake was also completed within 30 minutes. However, since the anoxic retention time was only 35 minutes on the B Side, there was little opportunity for post-denitrification (anaerobic) phosphorus release in the B Side anoxic zone (Figure 4b).

In the aerobic phase, phosphorus uptake in the A Side reactor to a bulk solution orthophosphate concentration of less than 0.1 mg P/L required approximately 120 minutes. Note that the rate of phosphorus uptake decayed in an exponential pattern across the aerobic phase in the A Side reactor (Figure 4a). On the other hand, because of the lower bulk solution phosphorus concentration at the end of the anoxic phase on the B side compared to the A Side, phosphorus uptake in the B Side reactor to a bulk solution orthophosphate concentration of less than 0.1 mg P/L required only 10 minutes of aeration, and the exponential decay in phosphorus uptake rate was not apparent (Figure 4b).

The results summarized in Figures 4a and 4b indicate that the reduction in anoxic HRT had a potentially beneficial effect on non-aerated phosphorus removal. The shorter anoxic zone HRT reduced the amount of phosphorus release that occurred after nitrate disappeared, resulting in a lower phosphorus loading to the aerobic phase, and less time required for the completion of phosphorus uptake.

As shown in Figure 4b, complete nitrification of ammonia by the suspended growth on the B Side took approximately 90 minutes, compared to only 10 minutes for complete phosphorus uptake. The suspended growth nitrifiers in the pilot plant mixed liquor were probably a mixture of fixed growth organisms sloughed from the FGR media, and bacteria which grew in suspension in the process mixed liquor. As described earlier, most of the nitrification in the pilot plant occurred in the FGRs,

upstream of the aeration basin. The results summarized in Figure 4b show that, by accomplishing some or all of the nitrification in the FGRs, the aeration basin size can be optimized for phosphorus removal only.
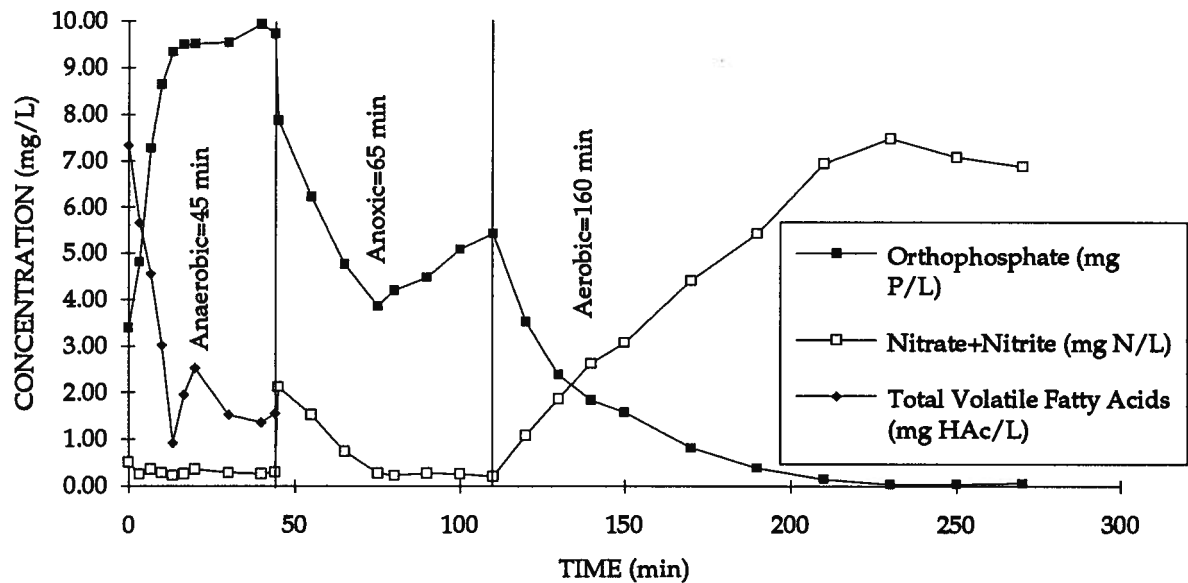

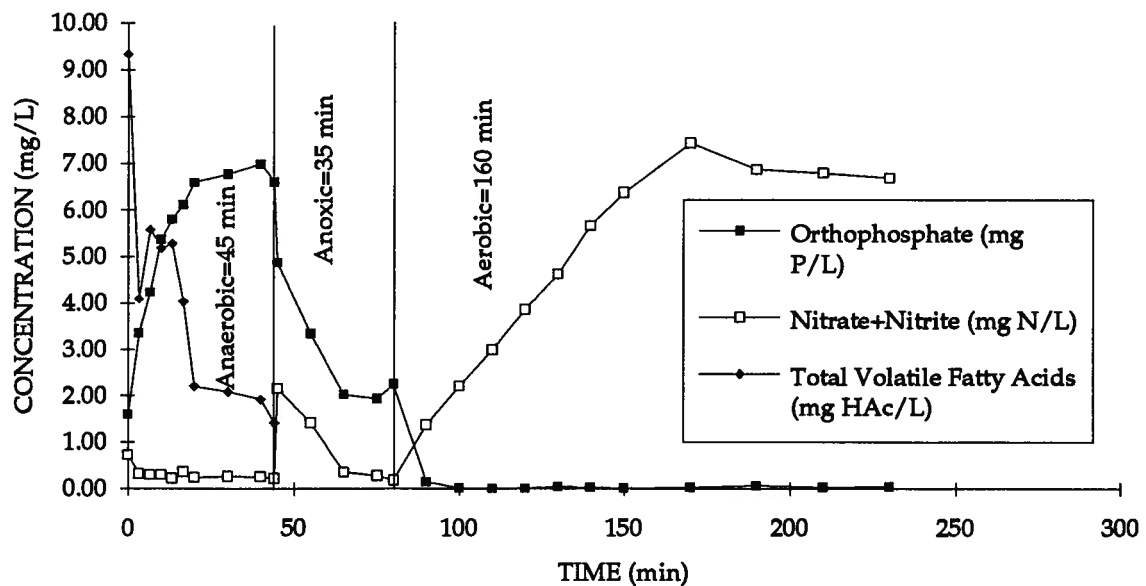Figure 4a - Phase 1 Batch Test - April 29, 1993 - A Side Process Train

Figure 4b - Phase 1 Batch Test - April 29, 1993 - B Side Process Train

The behavior in the anaerobic zone during Phase 1 was similar for both the A side (Figure 4a) and the B side (Figure 4b); that is, rapid VFA uptake with associated (rapid) phosphorus release was completed within the first 20 minutes for both A and B sides; after that point, VFA uptake-phosphorus release continued at a much slower rate, and there appeared to be little benefit in extending the anaerobic phase beyond the initial, rapid rate of phosphorus release-VFA uptake. Note that the rate of phosphorus release-VFA uptake slowed before all available VFA were removed from solution, and that the concentration of total VFA at the end of the anaerobic phase was 1-2 mg HAc/L for both the A and B Sides, indicating that bacterial polyP reserves were becoming exhausted before all available VFA could be stored.

The orthophosphate concentration at time=0 in the A Side batch reactor (approximately 3.4 mg P/L-see Figure 4a) was considerably greater than that in the B Side reactor (approximately 1.8 mg P/L-see Figure 4b). This can be explained by the apparent greater degree of secondary phosphorus release in the anoxic phase on the A Side; that is, since the orthophosphate concentration at the end of the anoxic phase in the A Side (approximately 5.5 mg P/L) was greater than on the B Side (approximately 2.5 mg P/L), the denitrified recycle from the anoxic reactor to the anaerobic reactor (see Figure 1) would carry a higher orthophosphate concentration on the A Side.

The other batch test simulation of Phase 1 SGR performance (results not shown) was conducted on March 31, 1993, during a period when phosphorus removal on both the A and B Side process trains was poor (see Figure 3). The poor pilot plant phosphorus removal was reflected in the March 31 batch test. However, the performance in the anoxic phase of the March 31 batch test confirmed that secondary phosphorus release began as soon as all nitrates were removed from solution, and that the lower anoxic HRT on the B Side resulted in a lower degree of post-denitrification phosphorus

release, compared to the A Side.

5.2.2 Phase 2 - 45 Minute Anaerobic Actual HRT vs. 25 Minute Anaerobic Actual HRT

The results of Phase 1 operation suggested that the size of the anaerobic reactor could be substantially reduced, and this was undertaken during Phase 2. For Phase 2, an anaerobic actual HRT of 45 minutes on the A Side was compared to one of 25 minutes on the B Side (Table 3). Two batch test simulations were conducted during Phase 2. The results of one of the Phase 2 batch tests (conducted on June 16, 1993) are summarized in Figures 5a (A Side) and 5b (B Side).

Note that the batch test shown in Figure 5 was conducted in June of 1993, shortly before it was discovered that excess solids were accumulating on the FGR media, and before the hydraulic pulse loading routine was begun. As described earlier, before pulse loading of the FGRs was introduced, solids accumulation was more severe on the A Side than on the B Side, and phosphorus removal was more erratic on the A Side (see Figure 3). Therefore, comparison of the two sides during Phase 2 was complicated by the uncontrolled variation introduced by the unequal solids accumulation on the FGR media.

In spite of the complications described above, the batch test results shown in Figure 5 indicate that the reduction in anaerobic HRT from 45 minutes to 25 minutes did not result in a deterioration in phosphorus removal. For both the A and B Sides, VFA uptake with associated rapid phosphorus release was completed within the first 15 minutes of the anaerobic phase, indicating that the 25 minute anaerobic HRT was more than adequate (Figure 5). Note that all VFA were removed from solution in the anaerobic phase on both the A and B Sides during the Phase 2 batch test, contrary to the results shown earlier in Figure 4 for the Phase 1 batch test.
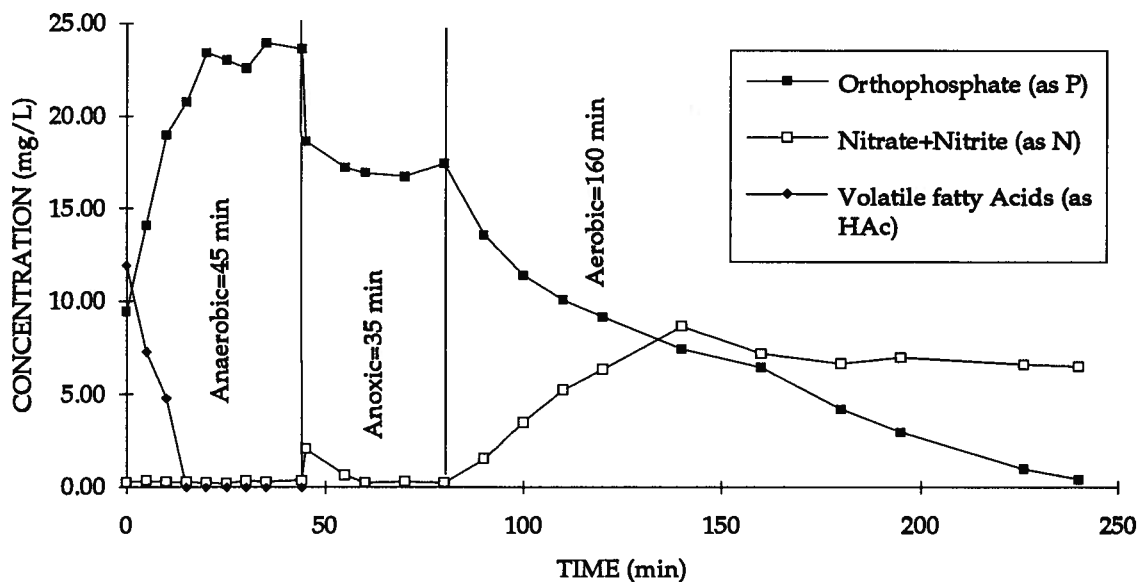
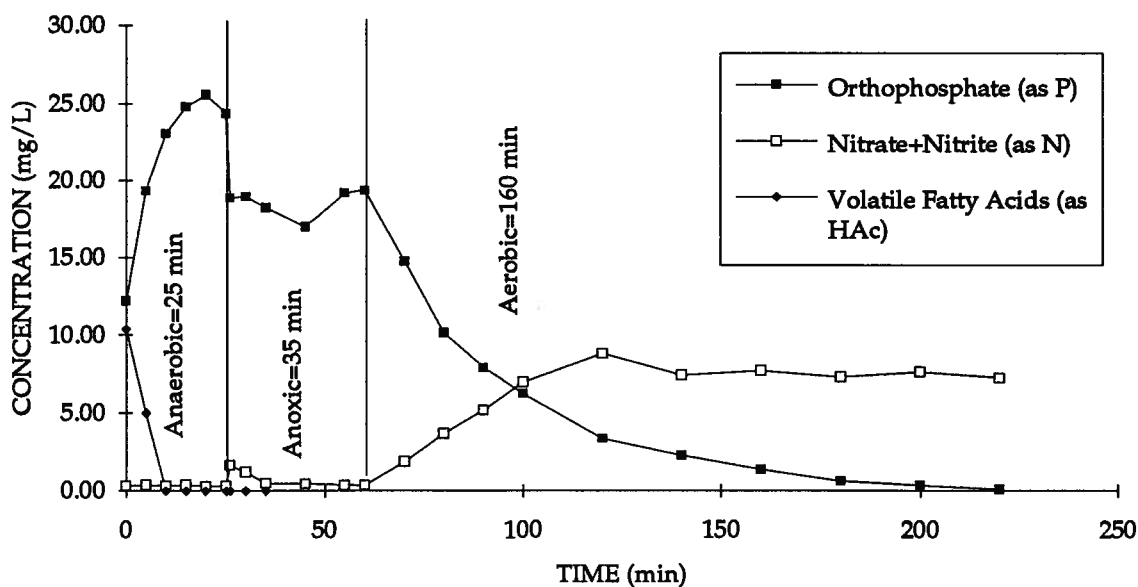Figure 5a - Phase 2 Batch Test - June 16, 1993 - A Side Process Train

Figure 5b - Phase 2 Batch Test - June 16, 1993 - B Side Process Train

Denitrification in the anoxic phase of the Phase 2 batch test on both the A and B Sides was accompanied by phosphorus uptake, with phosphorus release following the disappearance of nitrates from solution (Figure 5). The bulk solution orthophosphate

concentration at the end of the anoxic phase was approximately 20 mg P/L on both the A and B Sides. The exponential decay in aerobic phosphorus uptake rate observed in the A Side batch reactor in the Phase 1 batch test (Figure 4a) was apparent in both the A and B Side reactors for the Phase 2 batch test (Figures 5a and 5b, respectively). However, phosphorus removal to a bulk solution orthophosphate concentration of approximately 0.5 mg P/L took 160 minutes on the A Side, compared to only 120 minutes on the B Side (Figure 5). As shown earlier in Figure 3, phosphorus removal in the pilot plant to low effluent orthophosphate concentrations during June and July of 1993 was more consistent on the B Side than on the A Side, indicating that the above batch test results provided a reasonably accurate picture of pilot plant performance.

Contrary to the results of the Phase 1 batch test (Figure 4), nitrification of ammonia was completed before phosphorus uptake in both the A and B Side simulations for the Phase 2 batch test (Figure 5). Note that there was a much greater mass of orthophosphate released in the anaerobic phase of the Phase 2 batch test (approximately 15 mg P/L) than for the Phase 1 test (approximately 6 mg P/L); a longer aerobic period was consequently required for complete phosphorus uptake in the Phase 2 test. A second Phase 2 batch test was conducted on July 22, 1993, and the results (not shown) were similar to those shown above for the June 16 batch test.

5.2.3 Phase 3 - 25 Minute Actual Anaerobic HRT vs. 8 Minute Actual Anaerobic HRT

The results of Phase 2 operation (Figure 5) suggested that the anaerobic HRT could be further reduced, and this was undertaken during Phase 3. For Phase 3, the anaerobic volume on the A Side was reduced to give an actual HRT of 8 minutes, and the B Side anaerobic actual HRT was held at 25 minutes. The results of a typical batch test conducted during Phase 3 to compare the two process trains are shown in Figure 6.
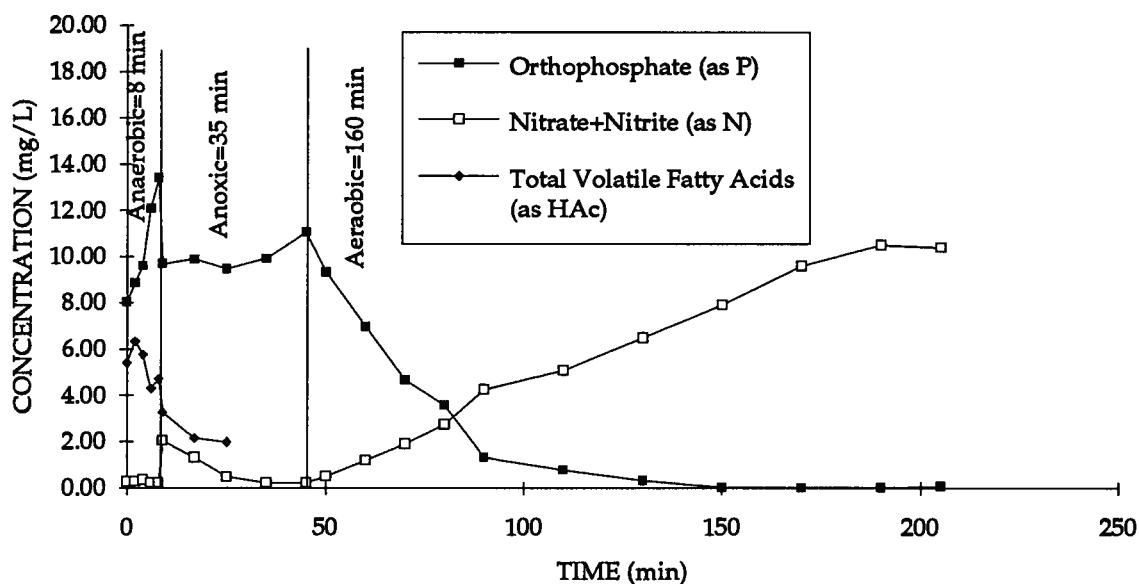
Figure 6a - Phase 3 Batch Test - Dec. 2, 1993 - A Side Process Train
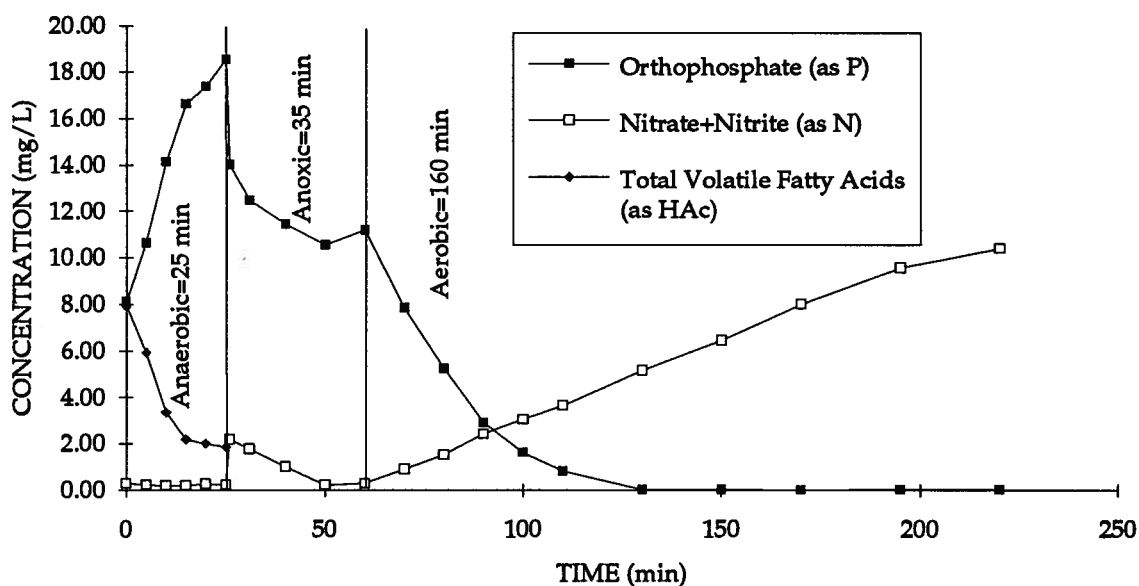
Figure 6b - Phase 3 Batch Test - Dec. 2, 1993 - B Side Process Train

The results the of Phase 3 batch tests (Figure 6) indicated that the reduction in anaerobic HRT from 25 minutes to 8 minutes on the A Side should not lead to an increase in effluent phosphorus concentration. On the A Side, VFA remained in solution at the end of the 8 minute anaerobic phase, and VFA removal in the subsequent anoxic phase was observed (Figure 6a). On the B Side, the rate of VFA

uptake with associated phosphorus release slowed at approximately t= 15 minutes, before all VFA were removed from solution, and the VFA concentration at the end of the anaerobic phase was approximately 2 mg HAc/L (Figure 6b), similar to the results shown in Figure 4 for the Phase 1 batch test. The mass uptake of VFA in the anaerobic phase of the Phase 3 batch test on the A Side was approximately 3 mg HAc/L, compared to 6 mg HAc/L on the B Side. Anaerobic orthophosphate release was correspondingly lower on the A Side (approximately 6 mg P/L), compared to the B Side (approximately 10 mg P/L).

The shorter anaerobic phase (and consequent lower mass of anaerobic phosphorus release-VFA uptake) on the A Side resulted in a lower phosphorus concentration at the outset of the anoxic phase on the A Side (approximately 10 mg P/L - Figure 6a) compared to the B Side (approximately 14 mg P/L - Figure 6b). However, the lower mass of anaerobic phosphorus release did not result in a lower phosphorus load on the aerated phase. Denitrification on the B Side was accompanied by phosphorus uptake (Figure 6b), similar to the results observed during the Phase 1 and 2 batch tests shown earlier in Figures 4 and 5, respectively.

On the other hand, denitrification on the A Side was not accompanied by a net phosphorus uptake (Figure 6a). Possibly, VFA available in solution at the outset of the anoxic phase on the A Side resulted in a continuation of VFA uptake by phosphorus accumulating bacteria with associated phosphorus release during the initial stage of the anoxic phase; the phosphorus release might tend to replace the phosphorus taken up during denitrification after the disappearance of VFA from solution, resulting in little or no net phosphorus uptake in the anoxic zone. The possible effects of high inputs of VFA to the anoxic phase are discussed later in this thesis.

In any case, the orthophosphate concentration at the end of the anoxic phase (resulting from the sum of anaerobic phosphorus release and anoxic phosphorus

uptake and/or release) was approximately 11 mg P/L for both the A and B Sides. The total HRT required for the anaerobic and anoxic phases, and for completion of aerobic phosphorus uptake, was approximately 130 minutes for both the A and B Sides. Again, an exponential decay in phosphorus uptake rate was observed across the aerobic phase in both the A and B Side reactors. The Phase 3 batch test results indicate that the shorter anaerobic HRT on the A Side (8 minutes) had neither a detrimental nor a beneficial effect on process performance, compared to the 25 minute HRT on the B Side. Based on results to this point, there appeared to be little benefit (or harm) in reducing the anaerobic HRT to 8 minutes, compared to 25 minutes.

Note that, similar to the batch test results described for Phase 1 (Figure 4), the batch tests conducted during Phase 3 (Figure 6) show that orthophosphate uptake on both the A and B Sides took significantly less time than nitrification of ammonia by the suspended growth. Three other batch tests conducted during Phase 3 (on December 16 of 1993, and on January 10 and February 7 of 1994) showed similar patterns to those shown above in Figure 6.

The results shown in Figures 4 and 6 indicated that the aerated actual SGR HRT could be substantially reduced from 160 minutes to 80-90 minutes, without compromising biological phosphorus removal. However, due to project time and budget constraints, the aerated HRT was not investigated.

5.2.4 Phase 4 - FGR Recycle Rate of 14 L/min vs. 24 L/min

In phase 4, an FGR recycle rate of 24 L/min on the A Side was compared to one of 14 L/min on the B Side; otherwise, all parameters between the two sides were identical (Table 3). It was not practical to attempt to simulate FGR performance in the batch tests, and parallel comparisons between FGR performance on the A and B Sides were restricted to pilot plant performance only. However, two batch test simulations of the

SGRs were conducted during Phase 4a, when the anaerobic actual HRT was 8 minutes on both sides (on May 2 and June 1 of 1994), and four were conducted during Phase 4b, when the anaerobic actual HRT on both sides was 25 minutes (on June 22, June 27, July 5, and July 11 of 1994). The purpose of the Phase 4 batch tests was to gain insight into the reasons for the deterioration in pilot plant phosphorus removal during that period. The results of one of the batch test simulations of the A and B Sides conducted during Phase 4a are summarized in Figure 7a and 7b, respectively.

The results shown in Figures 7a and 7b for the Phase 4a batch test in the anaerobic phase are similar to those shown for the A Side during Phase 3 (Figure 6a); that is, the bulk solution VFA concentration at the end of the anaerobic phase on both the A and B Sides was approximately 3 mg HAc/L, and the residual VFA was subsequently removed in the anoxic phase. Phosphorus release and denitrification occurred during the first 10 minutes of the anoxic phase; thereafter, denitrification was accompanied by phosphorus uptake. In the aerated phase, phosphorus uptake was weak on both sides; after 160 minutes of aeration, the bulk solution orthophosphate concentration was approximately 4 mg P/L on the A Side (Figure 7a), and 3 mg P/L on the B Side (Figure 7b). Again, the exponential decay in phosphorus uptake rate was observed in the aerobic phase in both the A and B Side reactors. It is apparent from the slope of the phosphorus uptake curve at the end of the batch test for both sides that additional aerobic HRT would not have substantially increased phosphorus uptake. The results shown in Figure 7 are consistent with pilot plant performance during Phase 4a; that is, plant effluent orthophosphate concentration was typically in the range 2-4 mg P/L, during June and July of 1994 (Figure 3).

In the batch tests conducted during Phase 4b, when the pilot plant anaerobic actual HRT was expanded to 25 minutes, complete or nearly complete uptake of all bulk solution VFA in the anaerobic phase was observed for both the A and B Sides in most

of the four tests, and phosphorus removal in both the batch tests and the pilot plant began to improve (Figure 3).
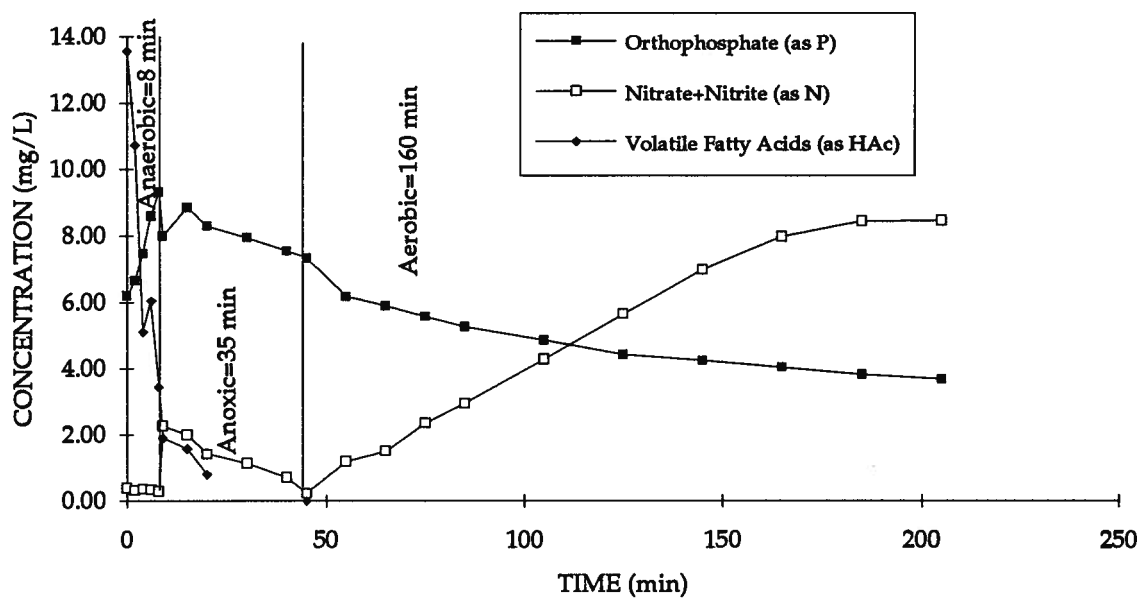

Figure 7a - Phase 4a Batch Test - June 1, 1994 - A Side Process Train
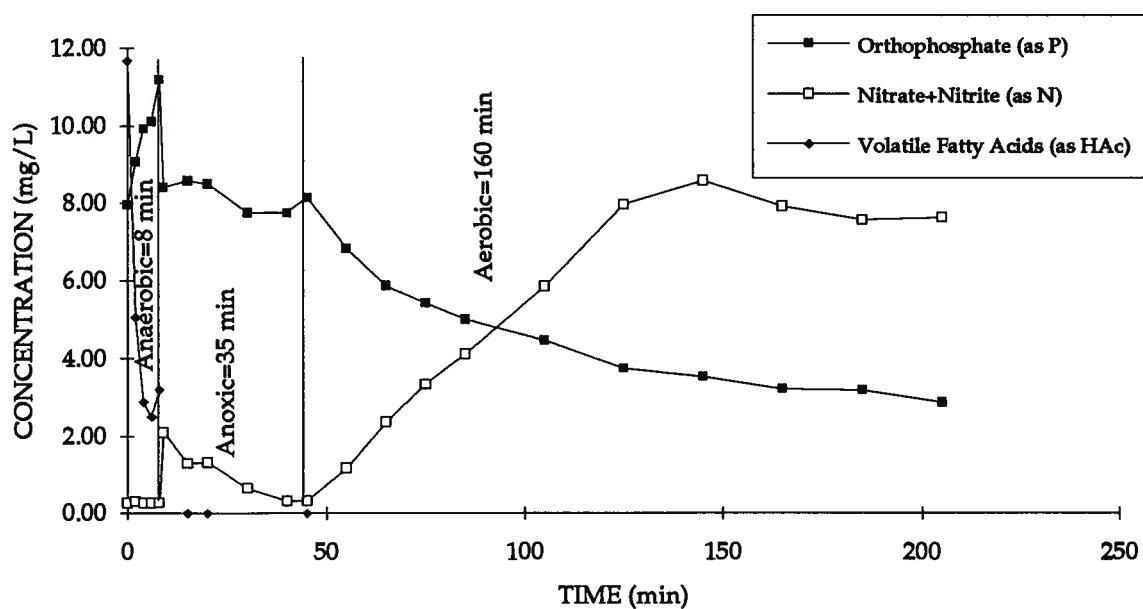

Figure 7b - Phase 4a Batch Test - June 1, 1994 - B Side Process Train

5.2.5 Phase 5 - Effects of Operating MLSS Concentration

During Phase 5, both the A and B Sides were configured identically, and the experimental variable was an aeration basin operating MLSS concentration of approximately 3,000 mg/L on the A Side, compared to 2,000 mg/L on the B Side. As noted earlier, there was no acclimation period between Phases 4 and 5, and the Phase 5 batch test results should be interpreted with this in mind. Two batch test simulations were conducted during Phase 5. The results of one of the tests are summarized in Figure 8. For the A Side (Figure 8a), orthophosphate in the anaerobic phase was released up to a bulk solution concentration of approximately 22 mg P/L, compared to only 12 mg P/L on the B Side (Figure 8b). On both sides, all available VFA were removed from solution before the end of the anaerobic phase; however, complete removal of VFA took only 10 minutes on the A Side (Figure 8a), compared to 20 minutes on the B Side (Figure 8b).

In the aerobic phase, the final orthophosphate concentration on the A Side was 0.65 mg P/L (Figure 8a), compared to 2.3 mg P/L on the B Side (Figure 8b). An exponential decay in aerobic phosphorus uptake rate was again observed in both the A and B Side reactors. In the B Side reactor, the slope of the curve at time=220 minutes indicates that additional aerobic HRT would have resulted in little further phosphorus uptake.

Note that the rate of denitrification in the anoxic phase was higher on the A Side (Figure 8a) than on the B Side (Figure 8b); denitrification in the A Side anoxic phase was complete within 20 minutes, while the B Side anoxic HRT of 35 minutes was barely adequate for complete denitrification. Similarly, the rate of nitrification by the suspended bacteria in the aerobic phase was faster on the A Side. The aerobic HRT of 160 minutes was adequate for complete nitrification of ammonia on the A Side, compared to the B Side, where more than 2 mg N/L of ammonia were left in solution at the end of the aerobic phase.
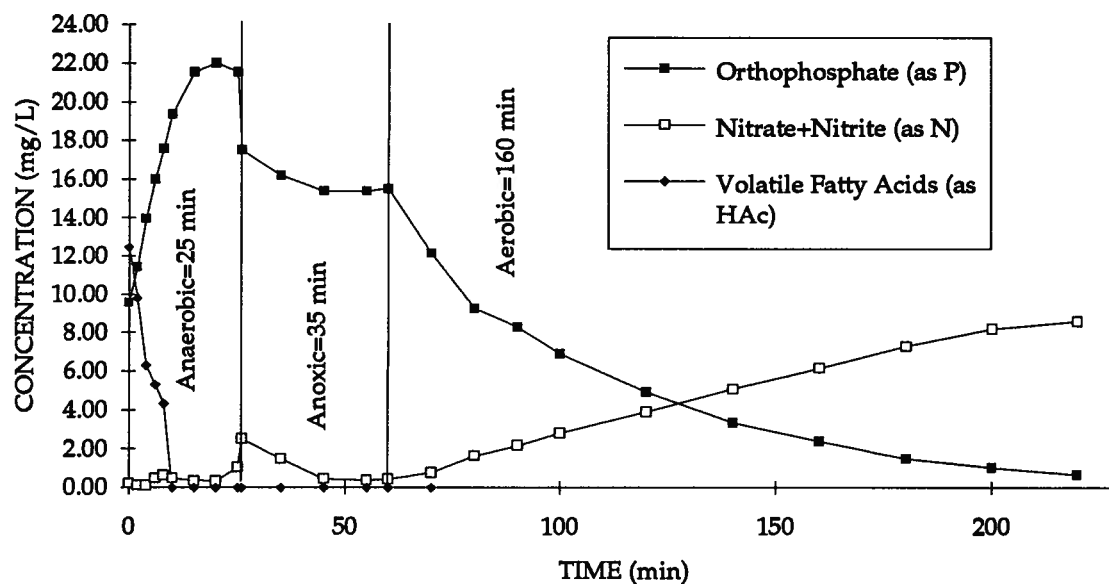
Figure 8a Phase 5 Batch Test - A Side Process Train - August 25, 1994
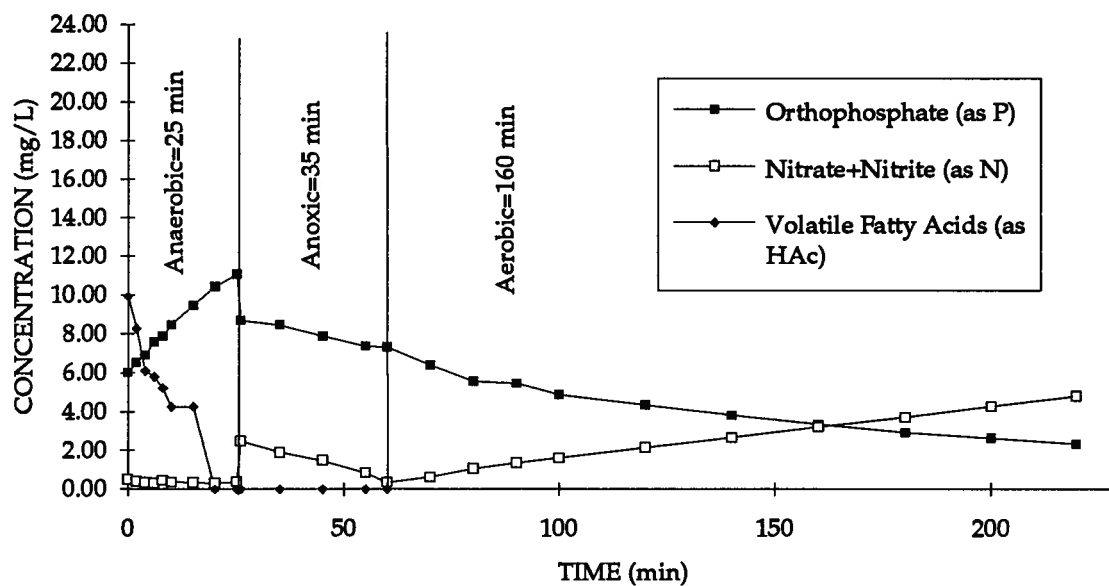
Figure 8b Phase 5 Batch Test - B Side Process Train - August 25, 1994

In light of the above, it is apparent that the lower sludge wasting rate on the A Side (which resulted in a mean aeration basin MLSS concentration of 3090 mg/L) resulted in more effective phosphorus removal and a greater suspended growth nitrification rate than the higher sludge wasting rate on the B Side (which resulted in an aeration

basin mean MLSS concentration of 1970 mg/L). The end result was a significant deterioration in biological phosphorus removal on the B Side, compared to the A Side. The batch test results for Phase 5 are supported by the results of pilot plant monitoring presented earlier in Figure 3 and Tables 13 and 14.

5.2.6 Simulation of Manipulating Denitrified Recycle Rate

As described in Chapter 4, the batch test simulation of manipulating the denitrified recycle rate was designed to investigate the single-cycle effects of increasing and decreasing the denitrified recycle flow rate in response to a simulated increase in process influent flow rate. The purpose of increasing or decreasing the denitrified recycle rate was to gain additional anaerobic actual HRT and to attenuate the dilution of anaerobic MLSS concentration, respectively. The simulated increase in process influent flow rate was 1.65 times the steady-state pilot plant influent flow rate (i.e., 1.65Q). Simulated decreases in the denitrified recycle rate to 1.5Q and Q and increases to 2.5Q and 3Q were compared to a simulation of the plant steady-state denitrified recycle rate of 2Q.

As noted in Chapter 4, the batch test simulated the reactor MLSS concentrations which would result only after the simulated flow rates had continued for a sustained period. In a full-scale situation, the change in reactor MLSS concentration resulting from a change in the steady-state recycle flow rate would happen gradually, and might not reach the steady-state value before the recycle rate was returned to its steady-state value after the peak flow was over. The batch test was designed only to give an indication of the effects of varying the denitrified recycle rate to attenuate the effects of plant influent flow fluctuations, and not to produce design data. However, the results provide some insight into the potential for increasing or decreasing the denitrified recycle to maximize uptake of VFA in the anaerobic reactor during daily peak plant loadings.

The results of the batch test investigation of denitrified recycle rate are summarized in Figures 9a-9e. Note that manipulation of the denitrified recycle flow rate in the pilot plant would affect the actual HRT and MLSS concentration in the anaerobic and anoxic reactors only (See Figure 1).

For the control reactor (#1-Figure 9a), the simulated denitrified recycle rate was 2Q. As shown, the simulated increase in process influent flow rate from Q to 1.65 Q caused a reduction in anaerobic actual HRT to 22 minutes, compared to the process steady-state value of 25 minutes. The anaerobic actual HRT of 22 minutes was just long enough for complete removal of all volatile fatty acids (VFA) from solution, with associated rapid orthophosphate release.
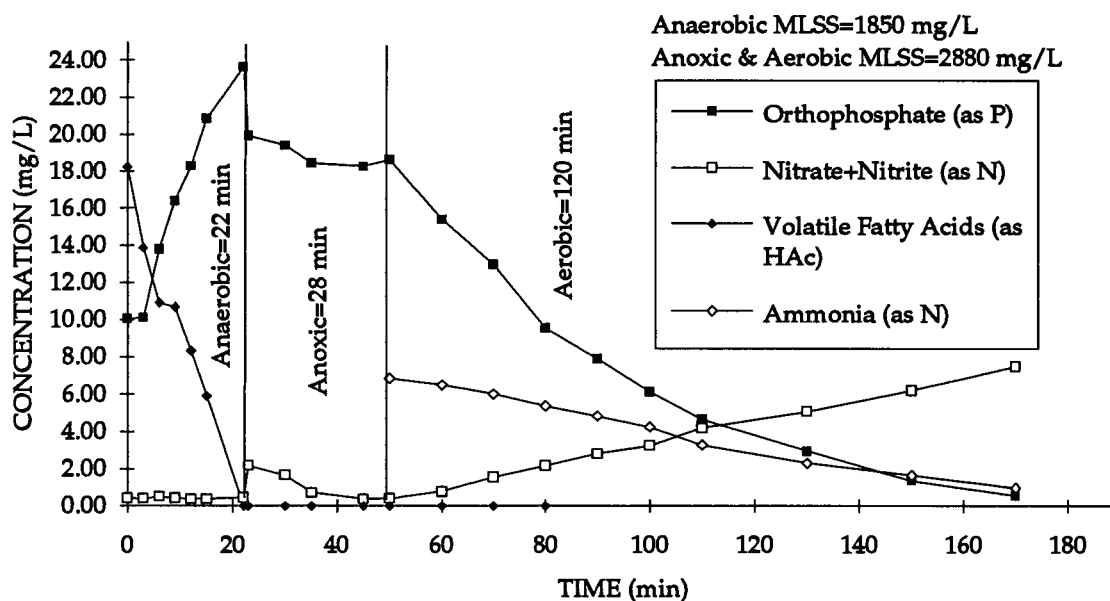
Figure 9a - Batch Reactor#1-Simulated Denitrified Recycle=2Q, August 23, 1994

The anoxic actual HRT of 28 minutes in the control reactor was more than adequate for complete denitrification of all available NO_x, with associated orthophosphate uptake (Figure 9a). The aerated actual HRT of 120 minutes was adequate for bacterial uptake-storage of orthophosphate to a bulk solution concentration of 0.55 mg P/L. Therefore, it appears that the pilot plant as it was operating on that day (August 23,

1994) had enough reserve capacity to handle a daily peak flow of 1.65Q, at least over a single cycle. Over a large number of regularly repeated cycles, performance might improve as the system biomass became acclimated to the daily peak loads.

The effects of simulating a reduction in the denitrified recycle from 2Q to 1.5Q (Reactor #2) are shown in Figure 9b. The decrease in recycle resulted in an increase in anaerobic actual HRT from 22 minutes to 25 minutes. However, due to a greater proportion of process influent added to the anaerobic phase, the decrease in recycle also resulted in a decrease in anaerobic MLSS concentration to 1680 mg/L (Figure 9b), compared to 1850 mg/L in the control reactor, and an increase in initial VFA concentration from 18 mg HAc/L in the control reactor to 20 mg HAc/L in Reactor #2 (Figure 9a). The net effect was that the concentration of VFA in solution at the end of the anaerobic phase in Reactor #2 was greater than 4 mg/L. As discussed earlier in the literature review, greater anaerobic storage of carbon (e.g., VFA) by bacteria should result in greater potential for phosphate uptake in the subsequent aerated phase. It follows that it is desirable to maximize bacterial uptake and storage of VFA in the anaerobic phase. Therefore, the situation illustrated in Figure 9b (i.e., VFA remaining in solution at the end of the anaerobic phase) should be avoided where possible. Over the long term, failure to allow maximum bacterial storage of VFA in the anaerobic phase might lead to a deterioration in biological phosphorus removal, at least during periods of peak phosphorus loadings.

The behavior in the anoxic phase for Reactor #2 was similar to that observed in Reactor #1; all available nitrate was removed, with associated orthophosphate uptake (Figure 9b). Unfortunately, the grab samples for orthophosphate from Reactor #2 for t=155 minutes and t=175 minutes were lost, so the end result of a simulated denitrified recycle of 1.5Q on phosphorus removal is unknown. However, it appears from the slope of the orthophosphate uptake curve to t=135 minutes that the final

orthophosphate concentration might have been well under 1 mg P/L at t=175 minutes (Figure 9b), indicating similar phosphorus removal effectiveness to Reactor #1.

Figure 9b-Batch Reactor #2-Simulated Denitrified Recycle=1.5Q, August 23, 1994

The effects of simulating a further reduction in the denitrified recycle to Q (Reactor #3) are shown in Figure 9c. In this case, the decrease in recycle increased the anaerobic HRT to 30 minutes, but reduced the anaerobic MLSS concentration to 1320 mg/L, and increased the initial VFA concentration to 22 mg HAc/L. Similar to Reactor #2, the net result was incomplete bacterial uptake-storage of VFA in the anaerobic phase. For Reactor #3, the VFA concentration at the end of the anaerobic phase was more than 8 mg/L, compared to 4 mg/L in Reactor #2.

Contrary to Reactors #1 and #2, orthophosphate release was observed during the initial 5 minutes of the anoxic phase in Reactor #3; at the same time, the remaining VFA disappeared from solution (Figure 9c). After the VFA concentration in the anoxic phase reached zero, phosphate uptake was observed until all available NO_x was denitrified. At the end of the aerated phase, the bulk solution orthophosphate concentration was 1.7 mg P/L, substantially higher than in Reactor #1.

Figure 9c -Batch Reactor #3 - Simulated Denitrified Recycle=Q, August 23, 1994

The effects of simulating an increase in the denitrified recycle to 2.5Q (Reactor #4) are shown in Figure 9d. In this case, the anaerobic MLSS concentration was 2150 mg/L (Reactor #4-Figure 9d), compared to 1850 mg/L in the control reactor (#1 - Figure 9a), the actual anaerobic HRT was reduced to 19 minutes, compared to 22 minutes in the control, and the initial VFA concentration was 12 mg HAc/L, compared to 18 mg HAc/L in the control. The net effect of increasing the denitrified recycle was that all available VFA were removed from solution within the first 12 minutes of the anaerobic phase. Increasing the denitrified recycle to 2.5Q reduced the actual anoxic HRT to 25 minutes, compared to 28 minutes in the control reactor; however, the shorter HRT was still adequate for complete denitrification (Figure 9d). Some orthophosphate uptake was observed during denitrification in the anoxic phase of Reactor #4. At the end of the aerated phase, the bulk solution orthophosphate concentration in Reactor #4 was 2.0 mg P/L, compared to only 0.55 mg P/L in the control reactor (#1).
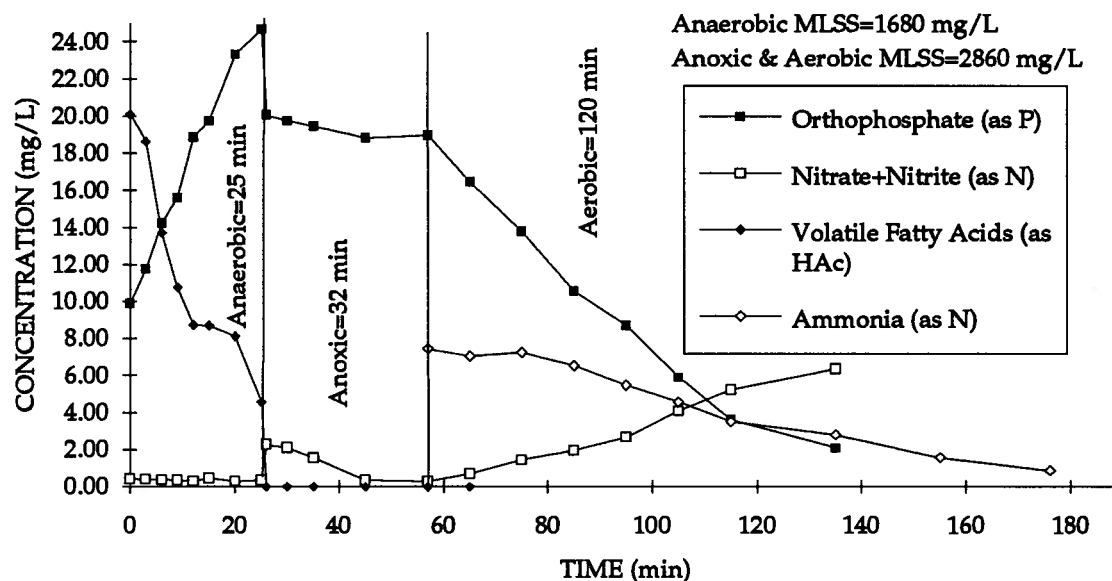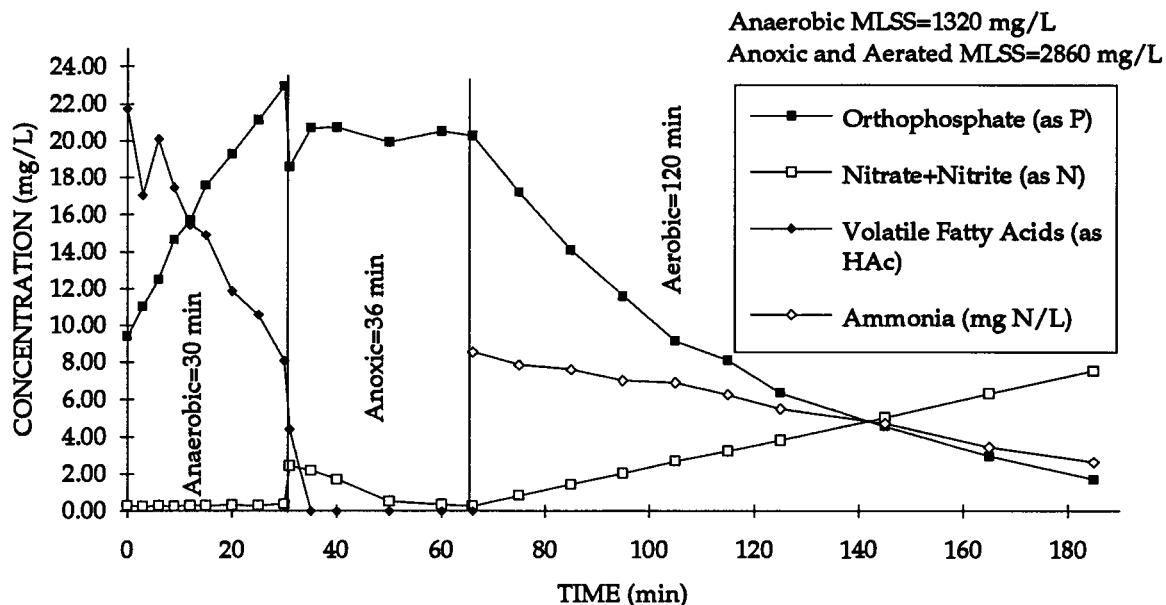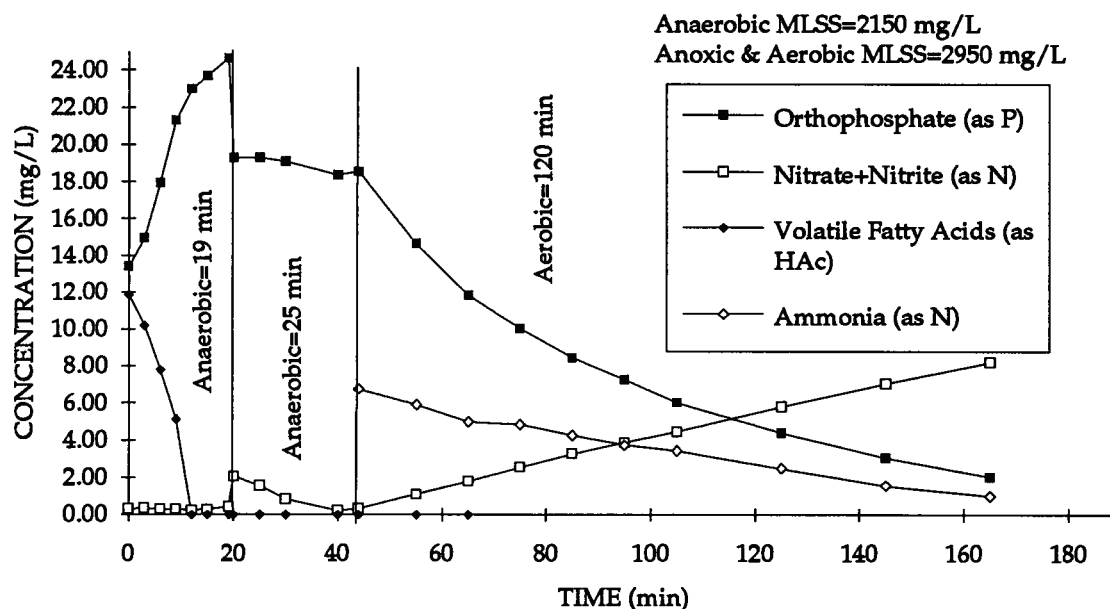
Figure 9d-Batch Reactor #4-Simulated Denitrified Recycle=2.5Q, August 23, 1994

The effects of simulating a further increase in the denitrified recycle to 3Q (Reactor #5) are shown in Figure 9e. The sample for anaerobic MLSS for Reactor #5 was lost. However, the value was estimated by mass balance calculations to be approximately 2200 mg/L, based on the MLSS concentration in the anoxic zone of the process, and dilution by the process influent. For Reactor #5, the simulated denitrified recycle flow of 3Q resulted in an actual anaerobic HRT of 17 minutes, compared to 22 minutes in the control, and an initial VFA concentration of 10 mg HAc/L, compared to 18 mg HAc/L in the control. All VFA in the process influent to the anaerobic phase of Reactor #5 were removed from solution within the first 9 minutes, compared to 12 minutes in Reactor #4. The anoxic actual HRT was reduced to 23 minutes by the increase in denitrified recycle, compared to 28 minutes in the control reactor; however, the 23 minutes was more than adequate for complete denitrification. Little phosphate uptake was observed during denitrification in the anoxic phase of Reactor #5, and the bulk solution orthophosphate concentration at the end of the aerobic phase was 1.1 mg P/L.
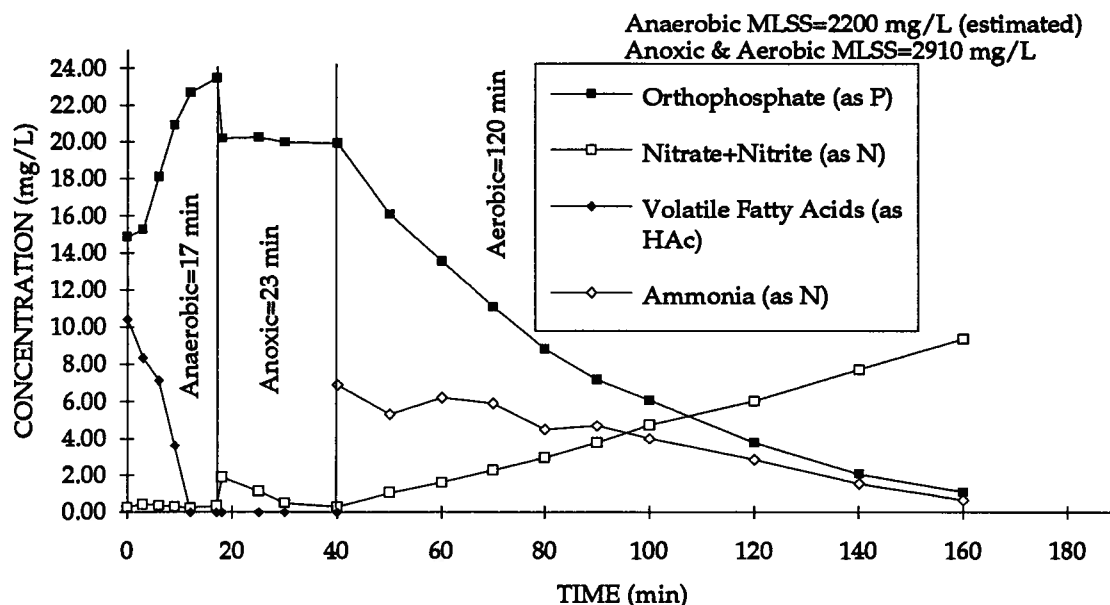
Figure 9e-Batch Reactor #5-Simulated Denitrified Recycle=3Q, August 23, 1994

In summary, for the batch test simulation of manipulating the denitrified recycle rate, the best phosphorus removal performance was observed in the control reactor, where the simulated recycle rate was equal to the pilot plant steady-state value (2Q), and the final orthophosphate concentration was 0.55 mg P/L. The low final orthophosphate concentration indicates that the pilot plant had enough capacity for phosphorus removal on the day of the test to handle an increase in process influent flow rate from Q to 1.65Q, at least for a short time. Simulating a decrease in denitrified recycle rate to 1.5Q and Q to gain additional anaerobic HRT reduced the mass of VFA taken up in the anaerobic phase from 18 mg HAc/L in the control reactor to 16 mg HAc/L and 14 mg HAc/L, respectively, due to increased dilution of the anaerobic MLSS concentration and increasing initial VFA concentration with decreasing denitrified recycle rate. Simulating an increase in the denitrified recycle to 2.5Q and 3Q to attenuate the dilution of anaerobic MLSS concentration resulted in a decrease in the anaerobic HRT required for uptake of all available bulk solution VFA from 22 minutes in the control reactor to 12 minutes and 9 minutes, respectively, due to a greater mass of solids and a lower initial VFA concentration in the anaerobic phase.

5.2.7 Simulation of Settled Biosolids Recycle Rate

As described in Chapter 4, the batch test simulation of manipulating the settled biosolids recycle rate was designed to investigate the single-cycle effects of increasing and decreasing the denitrified recycle flow rate in response to a simulated increase in process influent flow rate. The purpose of increasing or decreasing the denitrified recycle rate was to gain additional aerobic and anoxic actual HRT, and to attenuate the dilution of anoxic and aerobic MLSS concentration, respectively. The simulated increase in process influent flow rate was 1.65 times the steady-state pilot plant influent flow rate (i.e., $1.65Q$). Simulated decreases in the settled biosolids recycle rate $0.5Q$ and $0.25Q$ and increases to $1.5Q$ and $2Q$ were compared to a simulation of the plant steady-state settled biosolids recycle rate of Q.

As noted in Chapter 4, the batch test simulated the reactor MLSS concentrations which would result only after the simulated flow rates had continued for a sustained period, if the solids density of the secondary clarifier sludge blanket did not change. In a full-scale situation, the change in reactor MLSS concentration resulting from a change in the steady-state recycle flow rate would happen gradually, and might not reach the steady-state value before the recycle rate was returned to its original value after the peak flow was over. Further, the change in return biosolids flow rate would cause a change in the density if the secondary clarifier sludge blanket, resulting in a change in the MLSS concentration in the biosolids recycle flow stream. The batch test was designed only to give an indication of the effects of varying the biosolids recycle rate to attenuate the effects of plant influent flow fluctuations, and not to produce design data. However, the results provide some insight into the potential for increasing or decreasing the biosolids recycle to prevent an increase in plant phosphorus concentrations during peak plant loadings.

The results of the batch test investigation of settled biosolids recycle rate are

summarized in Figures 10a-10e. Note that manipulation of the settled biosolids flow rate in the pilot plant would affect the actual HRT and MLSS concentration in the anoxic and aeration reactors only. The anaerobic reactor would only be affected to the degree that a change in the anoxic MLSS concentration would cause a gradual change in the anaerobic MLSS concentration (see Figure 1). For the batch test simulation, all five reactors were configured and operated identically for the anaerobic phase.

For the control reactor, the simulated biosolids recycle rate was the pilot plant steady-state value of 2Q. In the control (Reactor #1-Figure 10a), the anoxic actual HRT of 28 minutes resulting from a simulated increase in process influent flow rate from Q to 1.65 Q was more than adequate for complete denitrification of all available nitrate; however, no associated net orthophosphate uptake was observed. Note that there was approximately 6 mg/L VFA remaining in solution at the end of the anaerobic phase. Removal of the remaining VFA during the first few minutes of the anoxic phase was accompanied by orthophosphate release (Figure 10a).
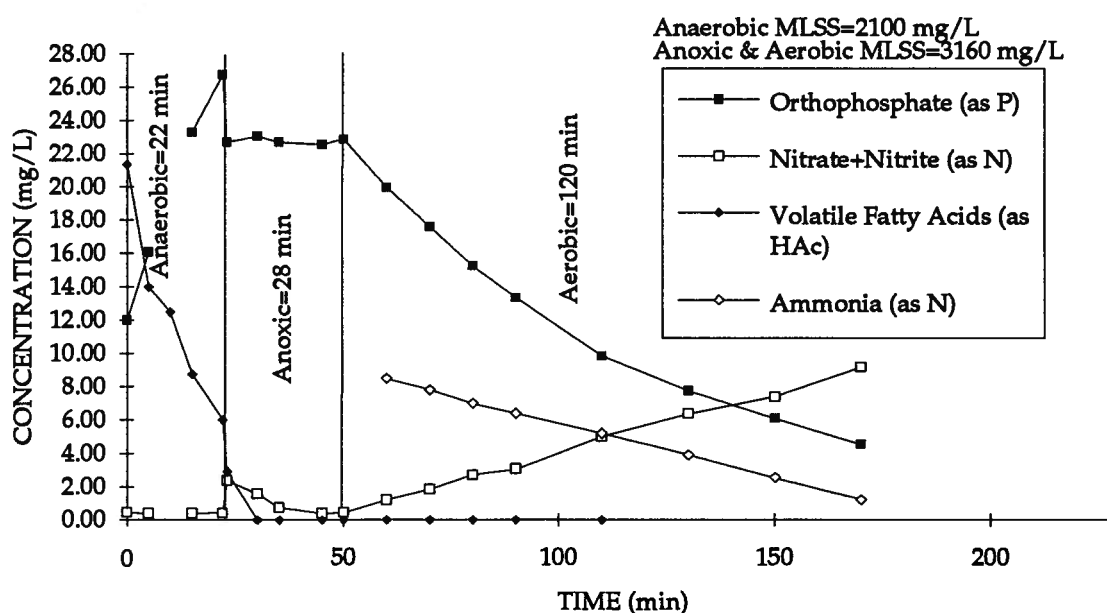
Figure 10a-Batch Reactor #1-Simulated Settled Biosolids Recycle=Q, August 29, 1994

The aerated actual HRT in the control reactor (120 minutes) was only long enough to result in bacterial uptake-storage of orthophosphate to a final concentration of 4.6 mg P/L (Figure 10a). Contrary to the results described earlier for the batch test investigation of denitrified recycle rate (Figure 10a), it appears that the pilot plant as it was operating on August 29, 1994, did not have enough reserve capacity to handle a daily peak flow of 1.65Q, at least over a single cycle.

The effects of simulating a reduction in the biosolids recycle from Q to 0.5Q (Reactor #2) are shown in Figure 10b. The decrease in recycle resulted in an increase in anoxic actual HRT from 28 minutes to 32 minutes, a decrease in the anoxic initial NO_x concentration from 2.4 mg N/L to 1.9 mg N/L, and an increase in aerobic HRT from 120 minutes to 147 minutes, compared to the control. However, the decrease in recycle also resulted in a decrease in anoxic and aerobic MLSS concentration to 2550 mg/L (Figure 10b), compared to 3160 mg/L in the control reactor (Figure 10a). In a flow-through system, a reduction in the return settled biosolids flow rate would result in a increase in the density of the secondary clarifier sludge blanket, and the steady-state anoxic and aerobic MLSS concentrations would gradually return to their original values. However, the immediate short-term effect of a reduction in the return biosolids flow rate would be a reduction in the solids input to the anoxic reactor, resulting in a dilution of the MLSS concentration.

The net effect of the simulation carried out in Reactor #2 was that the concentration of orthophosphate in solution at the end of the aerobic phase was 4.0 mg P/L, compared to 4.6 mg P/L in the control reactor. In the anoxic phase for Reactor #2, all available nitrate was removed, with associated orthophosphate uptake. Note that in this case, no VFA were remaining in solution at the end of the anaerobic phase, and a net anoxic uptake of orthophosphate was observed.
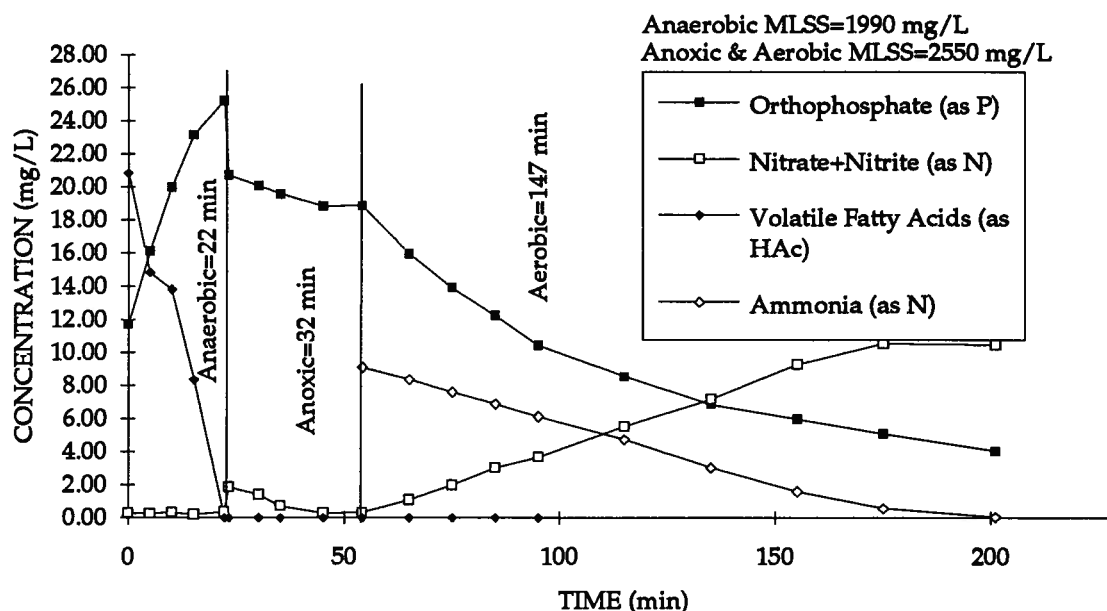
Figure 10b-Batch Reactor #2-Simulated Settled Biosolids Recycle=0.5Q, Aug. 29, 1994

The effects of simulating a further reduction in the biosolids recycle to 0.25Q (Reactor #3) are shown in Figure 10c. In this case, the simulated decrease in recycle increased the anoxic HRT to 34 minutes, decreased the anoxic initial NO_x concentration to 1.1 mg N/L, and increased the aerobic HRT to 166 minutes.
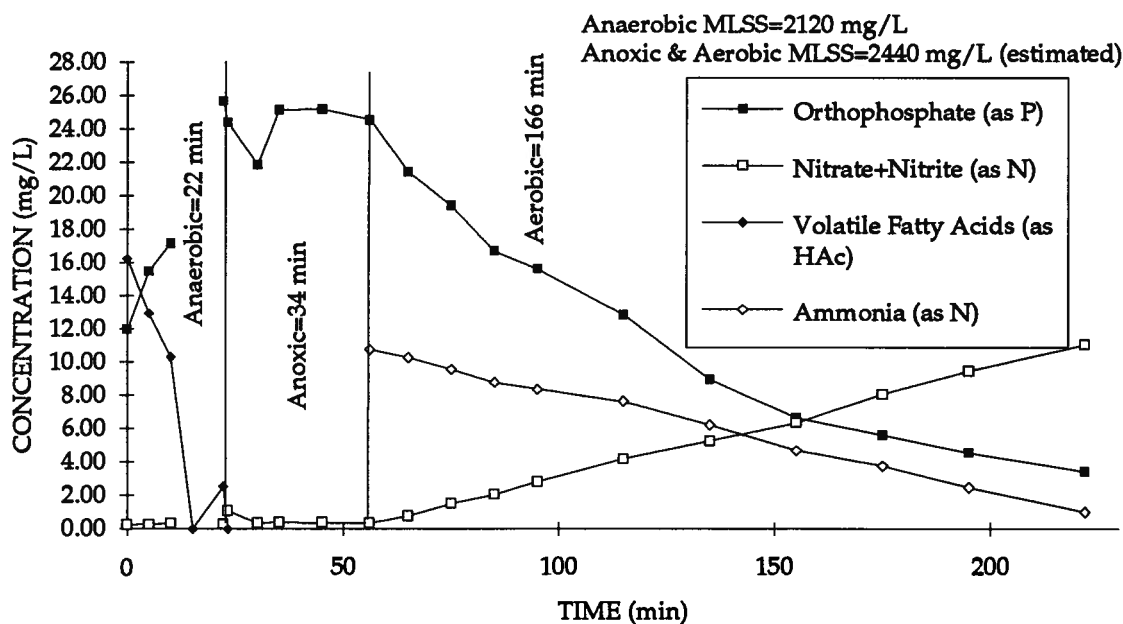
Figure 10c-Batch Reactor #3-Simulated Settled Biosolids Recycle=0.25Q, Aug. 29, 1994

The sample for anoxic and aerobic MLSS for Reactor #3 was lost. However, the value was estimated to be approximately 2440 mg/L, based on the MLSS concentration in the anaerobic zone of the process (Figure 10c). The data show approximately 2 mg/L VFA were remaining in solution at the end of the anaerobic phase in Reactor #3; however, since the preceding data point showed a non-detectable VFA concentration, these two data points appear to be unreliable (Figure 10c). No net orthophosphate uptake during denitrification in the anoxic phase was observed. At the end of the aerated phase, the bulk solution orthophosphate concentration was 3.4 mg P/L, compared to 4.0 mg P/L in Reactor #2 and 4.6 mg P/L in Reactor #1.

The effects of simulating an increase in the biosolids recycle to 1.5Q (Reactor #4) are shown in Figure 10d. In this case, the anoxic and aerobic MLSS concentration was increased to 3660 mg/L, the anoxic initial NO_x concentration was increased to 2.6 mg N/L, and the actual anoxic and aerobic HRTs were reduced to 25 minutes and 100 minutes, respectively. The anoxic HRT was sufficient for complete denitrification, and a net anoxic orthophosphate uptake was observed. At the end of the aerated phase, the bulk solution orthophosphate concentration in Reactor #4 was 4.1 mg P/L, similar to that observed in Reactor #2 (4.0 mg P/L). Approximately 5 mg/L VFA were remaining in solution at the end of the anaerobic phase in Reactor #4 (Figure 10d).

The effects of simulating a further increase in the biosolids recycle to 2Q (Reactor #5) are shown in Figure 10e. The anoxic and aerobic HRTs were reduced to 23 minutes and 87 minutes, respectively, the anoxic initial NO_x concentration was increased to 3.1 mg N/L, and the anoxic-aerobic MLSS concentration was increased to 3820 mg/L. Again, complete denitrification was observed in the anoxic phase, with associated orthophosphate uptake. The bulk solution orthophosphate concentration at the end of the aerobic phase was 3.3 mg P/L, similar to that observed in Reactor #3 (3.4 mg P/L). The concentration of VFA remaining in solution at the end of the anaerobic

phase in reactor #5 was approximately 5 mg HAc/L (Figure 10e).
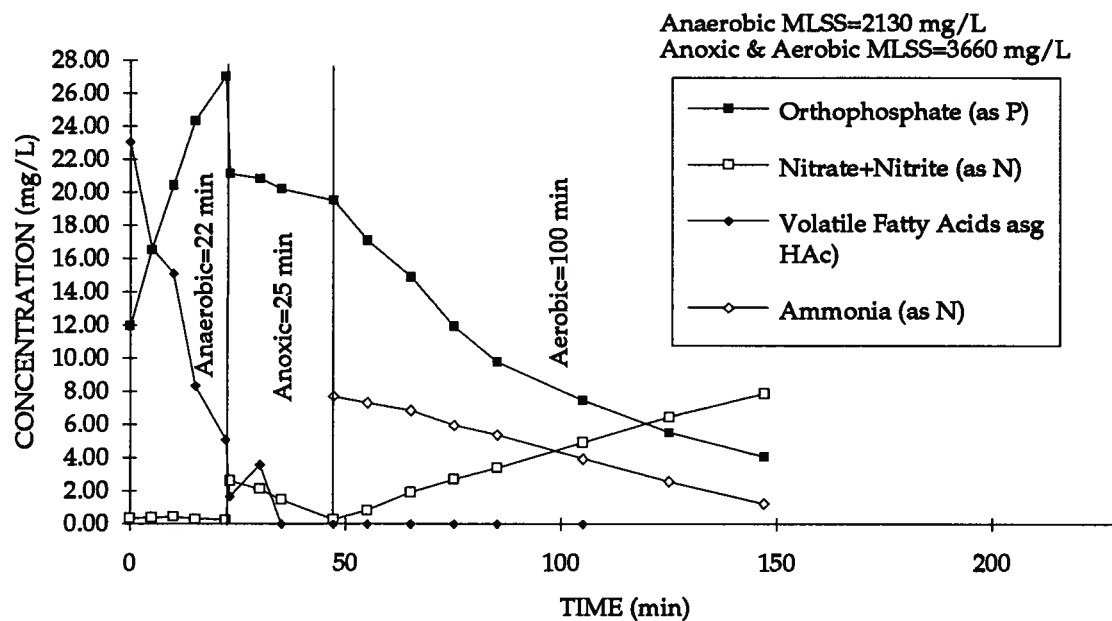
Figure 10d - Batch Reactor #4-Simulated Settled Biosolids Recycle=1.5Q, August 29, 1994
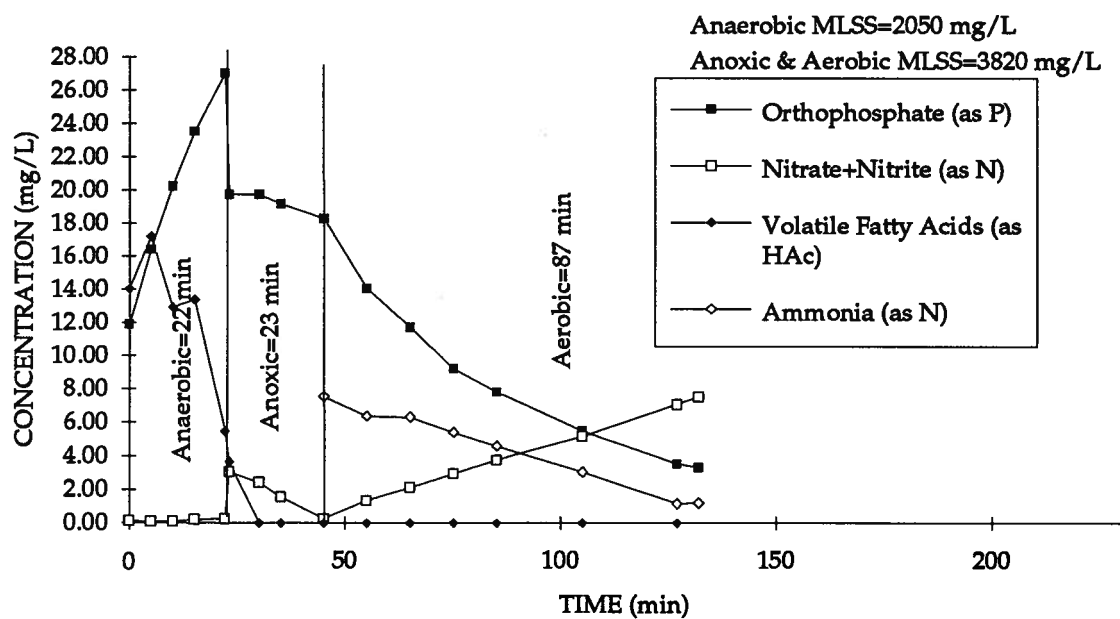
Figure 10e-Batch Reactor #5-Simulated Settled Biosolids Recycle=2Q, August 29, 1994

In summary, for the batch test simulation of manipulating the settled biosolids recycle flow rate, the poorest performance in terms of phosphorus removal was in the control reactor, where the simulated recycle was maintained at the pilot plant steady-state value (Q), and the final orthophosphate concentration was 4.6 mg P/L. The high final orthophosphate concentration indicates that the pilot plant biomass did not have enough reserve capacity on the day of the test to deal with a 65% increase in process influent flow rate. However, the system might become acclimated over the long term to a repeated daily peak load. Simulating a decrease in the biosolids recycle rate to 0.5Q and 0.25Q to gain additional aerobic HRT improved phosphorus removal to final orthophosphate concentrations of 4.0 mg P/L and 3.4 mg P/L, respectively. On the other hand, simulating an increase in biosolids recycle to 1.5Q and 2Q to attenuate the dilution of the aeration reactor MLSS concentration also improved phosphorus removal, to final concentrations of 4.1 mg P/L and 3.3 mg P/L, respectively. The batch test results indicate that manipulation of the settled biosolids recycle rate has potential as a process control parameter to help deal with the daily peak load at full-scale plants, both through decreasing the recycle to gain additional actual HRT, and through decreasing the recycle to attenuate dilution of the process MLSS. However, it should be noted that decreasing the biosolids recycle reduces the NO_x loading to the anoxic reactor, and can be expected to reduce net anoxic phosphorus removal, by reducing the amount of phosphorus uptake during denitrification, and by increasing the opportunity for post-denitrification phosphorus release.

6. DISCUSSION

6.1 BATCH TEST RESULTS AS AN INDICATOR OF PILOT PLANT PERFORMANCE

As described in Chapters 4 and 5, the batch tests were designed to gain additional insight into substrate removal rates in the process suspended growth reactors. However, conditions in the batch test reactors were not identical to those in the completely mixed flow-through process reactors. In the process reactors, some degree of imperfect mixing and short-circuiting was likely. In the batch test reactors, complete mixing was easy to achieve, and the sequential nature of the batch tests eliminated the possibility of short-circuiting. In the anaerobic phase, some air entrainment in the process reactors likely occurred due to the turbulence of mixing. In the batch test reactors, air entrainment during mixing was eliminated by maintaining a blanket of nitrogen gas over the liquid surface, although some air entrainment occurred when the mixed liquor was added to the reactor. Further, the concentration of an individual substrate in the batch tests was relatively high at the beginning of a particular phase, and then increased (or decreased) over the course of that phase, simulating a plug flow regime. In the process reactors, a relatively low steady-state substrate concentration was maintained. Therefore, if the substrate removal rate depended on the bulk solution concentration of that substrate (i.e., first order or higher with respect to concentration), the batch tests might give misleading results.

However, the batch test results showed that most of the substrate removal and production rates by the suspended growth were approximately linear across a wide range of concentrations (i.e., the rates of anaerobic VFA uptake and associated phosphorus release, the rates of anoxic NO_x reduction and associated phosphorus

uptake, and the rates of aerobic NH_4^+ removal and associated NO_x production). The exception was the rate of phosphorus uptake in the aerobic phase, which usually displayed an exponential decay with time. However, it is likely that the decay in phosphorus uptake rate was caused by a decline in intracellular PHA reserves, and not by the decline in the aerobic bulk solution phosphorus concentration (i.e., first order with respect to PHA reserves), since there was no characteristic bulk solution orthophosphate concentration at which the phosphorus uptake rate began to decline (see batch test results in Figures 4 to 10).

In general, review of the results presented in Chapter 5 shows that the batch tests gave a reasonably accurate picture of process performance. The bulk solution orthophosphate concentration at the end of the aerated phase of each of the batch tests was consistent with pilot plant effluent orthophosphate concentrations at the time of the batch test.

Since the results of pilot plant monitoring gave only the influent and effluent substrate concentrations of each reactor, the minimum HRT could not be determined in cases where the concentration of a particular substrate was at or near zero in the effluent from that reactor. On the other hand, the batch tests could be used to estimate the minimum actual HRT required for the completion of a particular biochemical reaction. For example, the batch tests conducted during the acclimation period before Phase 1 indicated that the actual HRT in the anoxic reactor of 65 minutes was approximately twice as long as necessary for complete denitrification. The subsequent reduction in the size of the B Side anoxic reactor to give an actual HRT of 35 minutes resulted in an increase of only 0.2 mg N/L in the mean anoxic reactor NO_x concentration, compared the A Side (see Phase 1 batch test results in Figures 4a and 4b and results of pilot plant monitoring in Table 11).

The batch test results summarized in Figures 4-10 were based on grab samples. For

pilot plant monitoring, the parameters for mass balances on individual reactors were also based on grab samples, and the parameters for process influent and effluent were based on 24-hour composite samples (except for VFA, which was based on grab samples). Therefore, for the most part, the results gave a "snapshot" of the behavior of the process biomass at a single moment in time, and relatively wide variations in the results of both the batch tests and pilot plant monitoring were commonly observed. However, the batch test results generally provided a reasonably accurate indicator of pilot plant performance, as discussed below.

The specific orthophosphate release, specific VFA uptake, and the bulk solution VFA concentration at the end of the anaerobic phase from all of the batch tests conducted throughout Phases 1-5 are compared to the mean values from the pilot plant anaerobic reactor in Table 16. Note that the specific orthophosphate release and specific VFA uptake shown in Table 16 were calculated on a mass per mass basis, to allow direct comparison between the results of pilot plant monitoring and the batch tests (i.e., the values were calculated by dividing the anaerobic reactor MLSS concentration into the overall mass of substrate uptake or release across the entire anaerobic phase for both the pilot plant reactors and the batch reactors). A direct comparison of specific substrate uptake or release rates across the anaerobic reactor between the batch tests and the pilot plant was not possible, for the reasons discussed below.

As described in Chapter 5, the specific VFA uptake rates and specific phosphorus release rates for the pilot plant anaerobic reactors shown in Tables 12 and 14 were calculated by dividing the specific mass of substrate uptake or release across the complete mix anaerobic reactor by the reactor actual HRT. However, the batch test results showed that rapid VFA uptake with associated phosphorus release was typically completed within a time significantly less than the available anaerobic HRT;

the specific substrate uptake and release rates calculated for the pilot plant anaerobic reactor are therefore likely to be lower than the actual biochemical reaction rates. On the other hand, the average specific VFA uptake and phosphorus release rates calculated from the batch test results (discussed later in this thesis) were based on the slopes of the VFA uptake and phosphorus release curves only to the point where rapid VFA uptake-phosphorus release had ceased. A direct comparison between specific substrate uptake and release rates between the pilot plant results and batch test results is therefore not possible. However, calculation of the specific substrate release or uptake on a mass per mass basis does allow direct comparisons between the two sets of results, and this method of comparison is used in the following discussion.

During Phase 1, when both the A and B Side anaerobic reactors had an actual HRT of 45 minutes, the mean values of pilot plant specific VFA release and specific orthophosphate uptake were significantly greater on the A Side (see Tables 12, 14, and 16). The difference is not apparent in the batch test results (Table 16); however, only 2 batch tests were conducted, compared to 12 pilot plant monitoring samples.

During Phases 2, 4a, and 4b, both sides had the same anaerobic actual HRT, and no differences in anaerobic behavior between the two sides were apparent in either the results of the batch tests or of pilot plant monitoring. During Phase 3, when the A Side anaerobic actual HRT was 8 minutes compared to 25 minutes on the B Side, the specific VFA release and specific orthophosphate uptake were significantly lower and the VFA concentration at the end of the anaerobic phase was significantly higher on the A Side in the pilot plant (Tables 12 and 14); this is reflected in the batch test results (Table 16).

Table 16 - Comparison of Batch Tests with Pilot Plant Monitoring - Anaerobic Phase

Study Phase	Data Source	Anaerobic Specific PO_4^{3-} Release (mg P/g MLSS)		Anaerobic Specific VFA Uptake (mg HAc/g MLSS)		VFA Concentration at End of Anaerobic Phase (mg HAc/L)	
		A Side	B Side	A Side	B Side	A Side	B Side
1	Batch 1[1]	4.4	6.2	No Data	No Data	No Data	No Data
	Batch 2[1]	4.1	2.6	3.8	4.2	2.3	1.4
	Process[2]	4.9	4.1	5.3	3.4	2.7	2.5
2	Batch 1[1]	6.4	3.3	5.4	2.8	0.0	0.0
	Batch 2[1]	4.5	4.5	4.6	4.5	0.5	0.4
	Process[2]	6.0	4.9	5.9	4.5	1.2	0.7
3	Batch 1[1]	2.6	5.0	0.3	2.9	4.7	1.7
	Batch 2[1]	1.6	4.7	4.2	4.5	3.7	0.0
	Batch 3[1]	0.8	4.5	0.6	3.8	12.2	0.8
	Batch 4[1]	1.6	6.2	1.0	4.5	6.6	0.8
	Process[2]	2.2	4.8	3.0	4.8	5.3	1.8
4a	Batch 1[1]	1.8	2.1	0.9	0.8	2.4	1.8
	Batch 2[1]	1.6	1.2	5.1	3.5	3.5	2.5
	Process[2]	3.0	2.8	3.2	5.1	4.5	6.9
4b	Batch 1[1]	3.5	1.8	2.4	2.0	1.4	0.3
	Batch 2[1]	2.8	3.6	4.0	3.1	3.3	0.0
	Batch 3[1]	3.7	3.6	4.0	3.3	0.0	0.0
	Batch 4[1]	2.8	3.9	4.9	3.2	0.4	0.0
	Process[2]	4.2	5.0	5.1	5.0	1.2	0.0
5	Batch 1[1]	5.1	3.7	6.9	4.1	0.0	3.1
	Batch 2[1]	5.0	3.2	5.2	6.3	0.0	0.0
	Process[2]	6.0	5.0	7.1	6.4	0.2	4.5

[1] Calculated from batch test results
[2] Mean value from results of pilot plant monitoring

During Phase 5, when the operating MLSS concentration was held at a higher value on the A Side compared to the B Side, the specific VFA uptake and orthophosphate release was greater and the final anaerobic VFA concentration was correspondingly lower on the A Side according to the results of pilot plant monitoring; the results of the first batch test conducted during Phase 5 are consistent with the results of pilot plant monitoring, but the results of the second test are not. In general, comparison of the results of the batch tests with those of pilot plant monitoring show that the batch tests were reasonably consistent with pilot plant performance in the anaerobic phase.

Each value shown for pilot plant monitoring was based on the mean of a relatively large number of samples, while only 2 to 4 batch tests were conducted during each phase; however, where the mean value of the pilot plant data showed that release, uptake, or concentration of a given substrate was greater on one side than the other, the batch test results usually showed the same pattern. (Table 16).

The specific mass of anoxic orthophosphate uptake and NO_x reduction and the specific mass of aerobic orthophosphate uptake from all of the batch tests conducted throughout Phases 1-5 are compared to the mean values from the corresponding pilot plant reactors in Table 17. As described above for the anaerobic reactor, the substrate uptake in the pilot plant reactors and in the batch test reactors shown in Table 17 was calculated on a mass per mass basis, to allow direct comparisons between the two sets of results (note that, similar to VFA uptake and phosphorus release in the anaerobic reactor, the batch tests showed that nitrate reduction and phosphorus uptake in the anoxic phase were typically completed within a time significantly less than the available HRT).

As shown in Table 17, some variation between the pilot plant data and batch test data is apparent. However, note that the lesser degree of anoxic specific orthophosphate uptake on the A side compared to the B Side in the pilot plant during Phase 3 and the low degree of anoxic specific orthophosphate uptake on both sides during Phase 4a was reflected in the batch tests.

The specific mass of NO_x reduction was relatively consistent between the batch tests and pilot plant monitoring throughout Phases 1-5 (Table 17). There was more variation in the specific mass of orthophosphate uptake in the aerobic phase, but in most cases, the batch test results were reasonably consistent with those of pilot plant monitoring, as far as comparisons between the two process trains is concerned.

Table 17 - Comparison of Batch Tests with Pilot Plant Monitoring - Anoxic and
Aerobic Phases

Study Phase	Data Source	Anoxic Specific PO_4^{3-} Uptake (mg P/g MLSS)		Anoxic Specific NO_x Reduction (mg N/g MLSS)		Aerobic Specific PO_4^{3-} Uptake (mg P/g MLSS)	
		A Side	B Side	A Side	B Side	A Side	B Side
1	Batch 1[1]	-0.3	0.3	0.6	0.7	3.0	3.2
	Batch 2[1]	1.2	1.0	0.9	0.7	2.7	0.8
	Process[2]	1.7	1.2	1.0	0.7	3.4	2.8
2	Batch 1[1]	0.4	0.1	0.6	0.2	5.6	3.8
	Batch 2[1]	1.2	1.0	0.7	0.7	2.7	4.6
	Process[2]	1.5	1.5	0.4	0.4	3.6	2.5
3	Batch 1[1]	-0.4	0.9	0.5	0.6	3.5	3.6
	Batch 2[1]	-0.2	1.3	0.5	0.7	2.2	3.0
	Batch 3[1]	-1.3	1.2	0.6	0.8	4.3	3.1
	Batch 4[1]	-0.2	1.3	0.8	0.7	2.4	4.4
	Process[2]	0.3	1.2	0.5	0.5	2.6	4.1
4a	Batch 1[1]	-0.3	0.0	0.6	0.6	1.8	1.7
	Batch 2[1]	0.2	0.1	0.7	0.4	1.3	1.9
	Process[2]	0.7	1.0	0.6	0.6	1.8	1.9
4b	Batch 1[1]	0.5	0.4	0.6	0.4	3.1	0.7
	Batch 2[1]	0.0	0.5	0.4	0.4	3.5	4.2
	Batch 3[1]	0.5	0.9	0.6	0.7	1.9	2.5
	Batch 4[1]	0.3	0.5	0.4	0.4	3.9	3.7
	Process[2]	0.8	0.7	0.7	0.5	1.8	1.9
5	Batch 1[1]	0.9	0.5	0.6	0.8	4.4	2.5
	Batch 2[1]	0.6	0.6	0.6	1.0	4.1	2.3
	Process[2]	0.7	0.9	0.7	1.0	4.9	3.5

[1] Calculated from batch test results
[2] Mean value from results of pilot plant monitoring

6.2 THE EFFECTS OF ANAEROBIC VFA UPTAKE ON PHOSPHORUS REMOVAL

6.2.1 Theoretical Factors Limiting Phosphorus Removal

To be consistent with current biochemical models (e.g., reviewed by Wentzel et al., 1991b), the amount of carbon taken up and converted to PHA by phosphorus accumulating bacteria in the anaerobic phase can be limited by the following three factors:

- The amount of VFA available in the anaerobic phase - If the bulk solution VFA is limiting, the result should be rapid bacterial uptake of all available VFA from solution, with associated rapid release of orthophosphate ions. The steady-state concentration of VFA in a completely mixed, flow through anaerobic reactor in this case should be near zero. As soon as bulk solution VFA become unavailable, the PHA storage-phosphorus release reaction should stop, unless VFA are produced by fermentation in the anaerobic reactor.

- The amount of intracellular polyP available to regenerate the ATP required for PHA synthesis - If polyP reserves are limiting, the result should be rapid uptake of VFA (with associated rapid phosphorus release) until the polyP available for ATP regeneration is exhausted, at which point VFA uptake-phosphorus release should cease. The steady-state concentration of VFA in the anaerobic reactor in this case will be greater than zero, unless both VFA and polyP are limiting.

- The retention time in the anaerobic reactor - If the anaerobic HRT is not sufficient to allow rapid VFA uptake-phosphorus release to continue until either bulk solution VFA or intracellular polyP become limiting, rapid VFA uptake-phosphorus release should continue until the end of the anaerobic phase. The steady-state concentration of VFA in the anaerobic reactor in this case might be greater than zero.

Similarly, in the aerobic reactor, the amount of phosphorus taken up and stored as polyP by the process bacteria can be limited by the following factors:

- The amount of intracellular PHA available - If PHA reserves are limiting, the result should be bacterial uptake and storage of available phosphorus until the PHA available for polyP formation is exhausted, at which point bacterial removal of phosphorus from the bulk solution will be limited to basic growth and

metabolic requirements, or to the slower use of external carbon sources to trigger the storage pathway. The net result might be significant concentrations of phosphorus in the process effluent.

- The amount of phosphorus available in the aerobic bulk solution - If PHA reserves are such that the phosphorus uptake-PHA degradation reaction continues until all of the phosphorus available in the bulk solution has been taken up and stored as polyP by the process bacteria, the net result should be low or non-detectable concentrations of soluble phosphorus in the process effluent.

- The retention time in the aerobic reactor - If the aerobic HRT is such that the mixed liquor leaves the aeration reactor before either PHA reserves are exhausted or all of the available bulk solution phosphorus has been taken up and stored as polyP by the process bacteria, the net result might be significant concentrations of phosphorus in the process effluent (this situation will not arise in an acclimated system that includes nitrification).

- The supply of electron acceptors to the aerobic (or anoxic) phase - If the supply of molecular oxygen (or possibly nitrates) to the aerobic reactor is low enough to limit the rate of the phosphorus uptake-PHA degradation reaction, the net result might be similar to that for the aerobic HRT; that is, the process liquid might leave the aeration reactor before bacterial uptake and storage of phosphorus can be completed.

Note that the batch test results showed that aerobic phosphorus uptake was generally not limited by HRT in the pilot plant; in cases where significant concentrations of orthophosphate remained in the bulk solution at the end of the aerated phase of the batch tests, the exponential decay in phosphorus uptake rate indicated that additional aerobic HRT would result in little further phosphorus uptake. Nor was it likely that

the rate of phosphorus uptake was limited by the supply of electron acceptors in the pilot plant; the dissolved oxygen concentration in the process mixed liquor in the FGR catchbasin sumps and in the aeration basin was typically at least 5-6 mg/L.

It can be seen from the theoretical limitations on phosphorus removal discussed above that the ratio of VFA to phosphorus in the process influent is critical to process effluent phosphorus concentrations. In a steady-state system with no HRT limitations, if the ratio of VFA to phosphorus were high enough, the process bacteria would accumulate sufficient PHA reserves under anaerobic conditions to take up and store all available soluble phosphorus under subsequent aerobic conditions (at least during steady-state operation), and process effluent phosphorus concentrations would be low. If the ratio of VFA to phosphorus declined, anaerobic PHA storage might be insufficient to drive aerobic uptake and storage of all available phosphorus, and process effluent soluble phosphorus concentrations would rise. If the anaerobic HRT were insufficient to allow the PHA storage-polyP degradation reaction to continue until either all bulk solution VFA were taken up and stored or bacterial polyP reserves were exhausted, process effluent soluble phosphorus concentrations would depend on the relative amounts of anaerobic PHA storage and process phosphorus loading.

Full-scale treatment plants typically operate in a pseudo steady-state condition, where flow and load variations follow a cyclical pattern, and random variations in process influent character are common. In the FGR-SGR pilot plant, the process influent flow rate for both the A and B Sides was always kept at a steady-state value of 4.8 L/min. Further, the process influent was stored in completely-mixed storage tanks with an average HRT of 36 hours; raw sewage was pumped from the trunk sewer to the storage tanks on a regular timed cycle in an attempt to obtain the most consistent process influent quality possible. However, the character of the process influent was

still subject to wide variations. The ratio of VFA to total phosphorus (VFA:P) in the process influent (primary clarifier overflow) for the A and B Sides throughout Phases 1-5 is summarized in Figure 11. As shown, the VFA:P ratio in the influent to both processes varied from less than 4:1 to more than 20:1. Some differences between the A and B Sides are apparent, but the general pattern is similar for both. Note that the concentration of VFA was based on grab samples, and the concentration of phosphorus was based on 24-hour composite samples. Since the VFA:P ratio followed a similar pattern for both sides, the side by side comparisons of the experimental treatments should be valid. However, an experimental condition which resulted in good phosphorus removal at a high influent VFA:P ratio might result in poor phosphorus removal at a lower influent VFA:P ratio.

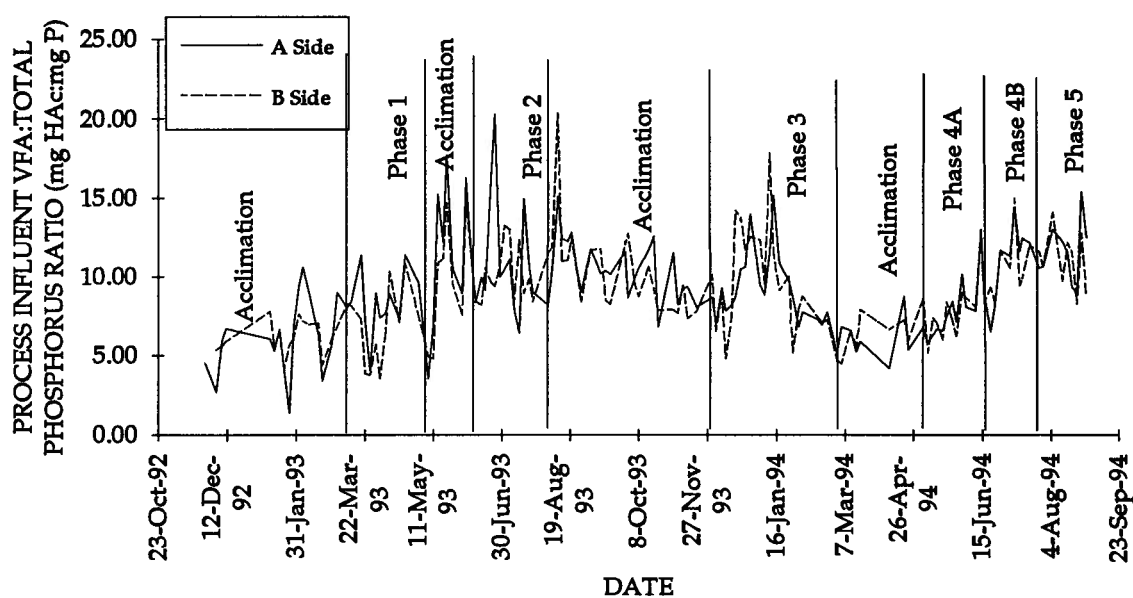Figure 11 - Process Influent Volatile Fatty Acid to Total Phosphorus Ratio

As discussed above, wide variations in the process influent VFA:P ratio could be expected to affect effluent phosphorus concentrations. Comparison of the data in Figure 11 to the plant effluent orthophosphate concentrations summarized earlier in Figure 3 shows that, except for the period of solids accumulation in the FGRs (August

to November of 1993), periods of relatively high influent VFA:P ratio are associated with low plant effluent phosphorus concentrations. The relationship between process influent VFA concentration, process influent total phosphorus concentration, and process effluent total phosphorus concentration is discussed in the following sections.

6.2.2 The Effects of Anaerobic VFA Uptake on Anaerobic Phosphorus Release and Aerobic Phosphorus Uptake

As described in the previous section, bacterial carbon storage and phosphorus release in the anaerobic phase are theoretically directly related. The linear rates of rapid VFA uptake and associated phosphorus release observed in the anaerobic phase of the batch tests (Figures 4-10) were compatible with the behavior of a typical BNR biomass, and the results of pilot plant monitoring were consistent with the batch test results (Table 16).

The results of pilot plant monitoring presented in Chapter 5 showed that periods of higher anaerobic VFA uptake (Table 12) were generally associated with periods of high anaerobic orthophosphate release (Table 14).The relationship between anaerobic VFA uptake and phosphorus release in the pilot plant is shown in Figure 12, where the specific release of orthophosphate in the anaerobic reactor of the pilot plant is plotted against the anaerobic specific uptake of VFA (as acetic acid). All of the data for both the A and B Sides throughout Phases 1-5, including the interim acclimation periods, are included in Figure 12. The specific VFA uptake and phosphorus release in Figure 12 are shown on a mass per mass basis; the use of specific uptake and release rates here would introduce a source of variation unrelated to the biochemical relationship, since the controlled variations in anaerobic HRT over the course of the study would result in large step changes in the calculated substrate uptake or release rate across the reactor (see earlier discussion in Section 6.1).

The data shown in Figure 12 confirm the direct relationship between orthophosphate

release and VFA uptake in the anaerobic phase. The slope of the regression line is 0.88 mg P released per mg VFA (as acetic acid) taken up. In terms of molar ratios, the slope of the regression gives 1.7 mole P released per mole VFA (as acetic acid) taken up, considerably higher than the theoretical 1:1 ratio predicted by Wentzel et al. (1991b). In bench scale batch tests of the biomass from the pilot-scale UCT process adjacent to the FGR-SGR pilot plant, Comeau et al. (1987) reported molar ratios of phosphorus release to VFA uptake in the range 1.5:1 to 1.7:1 in response to additions of mixtures of acetate and propionate. The full-scale study at the FGR-SGR plant in Salmon Arm showed that the molar ratios of phosphorus release to acetate uptake were in the range 1.1:1 to 1.8:1 (Gibb, 1990). The pilot plant data are reasonably consistent with the findings of Comeau et al. (1987) and Gibb (1990).
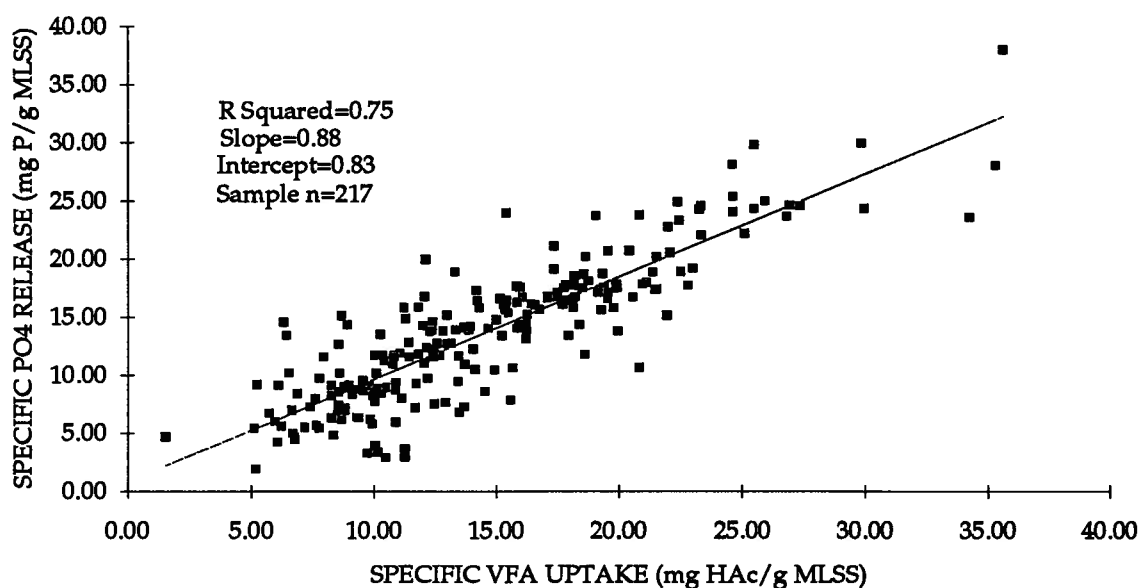
Figure 12 Pilot Plant Anaerobic Specific Phosphorus Release vs. Specific VFA Uptake - A and B Side Data Pooled

As described in the literature review and in Section 6.2.1, the amount of phosphorus taken up under aerobic conditions is theoretically directly related to the amount of

VFA uptake-phosphorus release in the preceding anaerobic phase. The results of pilot plant monitoring presented in Chapter 5 showed that periods of higher aerobic orthophosphate uptake were associated with periods of higher anaerobic orthophosphate release and anaerobic VFA uptake (Tables 12 and 14). The specific uptake of orthophosphate over the entire aerated phase of the pilot plant (i.e., the sum of the uptake across FGR Cells 1 and 2 and the aeration SGR) is plotted against the anaerobic specific orthophosphate release in Figure 13. Again, all of the data for both the A and B Sides throughout Phases 1-5, including the interim acclimation periods, are included. The data shown in Figure 13 confirm the direct relationship between anaerobic phosphorus release and aerobic phosphorus uptake.
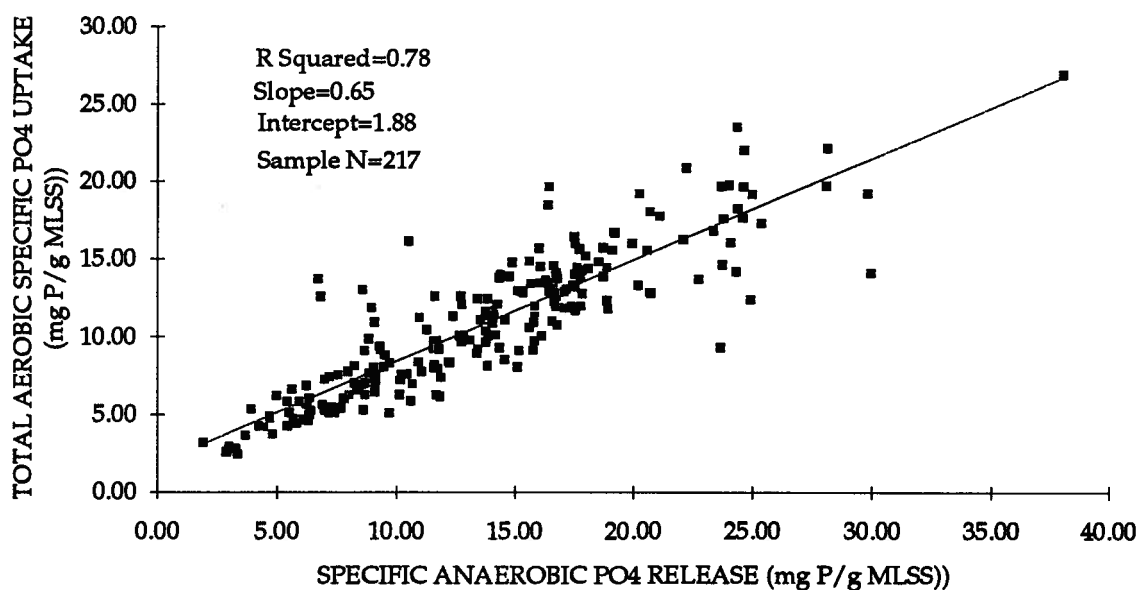

Figure 13 - Pilot Plant Aerobic Specific Phosphorus Uptake vs. Anaerobic Specific Phosphorus Release - A and B Side Data Pooled

The data summarized in Figures 12 and 13 show that the behavior of the pilot plant suspended biomass was consistent with current biochemical models. That is, the amount of phosphorus released in the anaerobic reactor was directly dependent on

the amount of VFA removed from solution, and the amount of phosphorus taken up in the subsequent aerobic phase was directly dependent on the amount of phosphorus released in the anaerobic phase. The earlier full-scale work at Salmon Arm also yielded linear relationships between VFA loading and phosphorus release in the anaerobic phase, and between anaerobic phosphorus release and aerobic phosphorus uptake (Gibb, 1990); however, direct comparisons with that data are not possible, since different units were used in the Salmon Arm correlations. It can be seen from the above that the amount of phosphorus taken up in the aeration reactor is directly related to the amount of VFA taken up in the anaerobic reactor. Phases 2 and 3 of the study involved manipulation of the anaerobic HRT, which could be expected to influence the amount of VFA taken up. The effects of the anaerobic HRT on phosphorus removal are discussed in the following section.

6.2.3 The Effects of Anaerobic HRT on Phosphorus Removal

As shown by the batch test results, as long as the actual HRT of the anaerobic reactor was 25 minutes or greater and the aeration basin operating MLSS concentration was approximately 3,000 mg/L (i.e., both the A and B Sides during Phases 1 and 2, the B Side during Phase 3, both sides during Phase 4b, and the A Side during Phase 5), rapid VFA uptake with associated phosphorus release was typically completed before the end of the anaerobic phase, indicating that plant phosphorus removal during those periods was not limited by anaerobic HRT (see Figures 4a, 4b, 5a, 5b, 6b, and 8a). As described in Chapter 5, an anaerobic actual HRT of 45 minutes on the A Side did not result in better phosphorus removal than an HRT of 25 minutes on the B Side (see the results of Phase 2 pilot plant monitoring in Section 5.1.6, and the Phase 1 batch test results in Section 5.2.2). In fact, the mean effluent orthophosphate concentration was significantly lower on the B Side, during Phase 2 (Table 13).

When the anaerobic actual HRT was 8 minutes (i.e., the A Side during Phase 3 and

both sides during Phase 4a), the batch test results showed that rapid VFA uptake with associated phosphorus release was typically not completed by the end of the anaerobic phase, indicating that plant phosphorus removal during those periods might have been limited by anaerobic HRT (see Figures 6a, 7a, and 7b). As described in Chapter 5, an anaerobic actual HRT of 8 minutes resulted in a significantly lesser degree of VFA uptake than did an HRT of 25 minutes (see the results of Phase 3 pilot plant monitoring in Section 5.1.5 and Phase 3 batch test results in Figure 6). However, the greater VFA uptake associated with the 25 minute HRT did not result in significantly better phosphorus removal compared to that associated with the 8 minute HRT during Phase 3, indicating that the carbon storage resulting from the 8 minute HRT on the A Side during Phase 3 was adequate for good phosphorus removal (see the results of Phase 3 pilot plant monitoring in Section 5.1.6 and Phase 3 batch test results in Figure 6). The deterioration in phosphorus removal when both sides were operated with an anaerobic actual HRT of 8 minutes following the completion of Phase 3 indicated that plant phosphorus removal might have become limited by anaerobic HRT at that point (see the results of pilot plant monitoring in Figures 3a and 3b, and the Phase 4a batch test results in Figure 7).

As described earlier (Section 6.2.1), in cases where the degree of rapid VFA uptake-phosphorus release is limited by HRT rather than by the amount of VFA available or by the amount of intracellular polyP reserves, the phosphorus concentration in the process effluent theoretically depends on the relative amounts of VFA taken up and stored as PHA and the process phosphorus loading. The ratio of VFA taken up in the anaerobic reactor to process influent total phosphorus concentration for the A Side throughout Phases 1-5 is shown in Figure 14. The A Side plant effluent orthophosphate concentration for the same period is summarized in Figure 15.
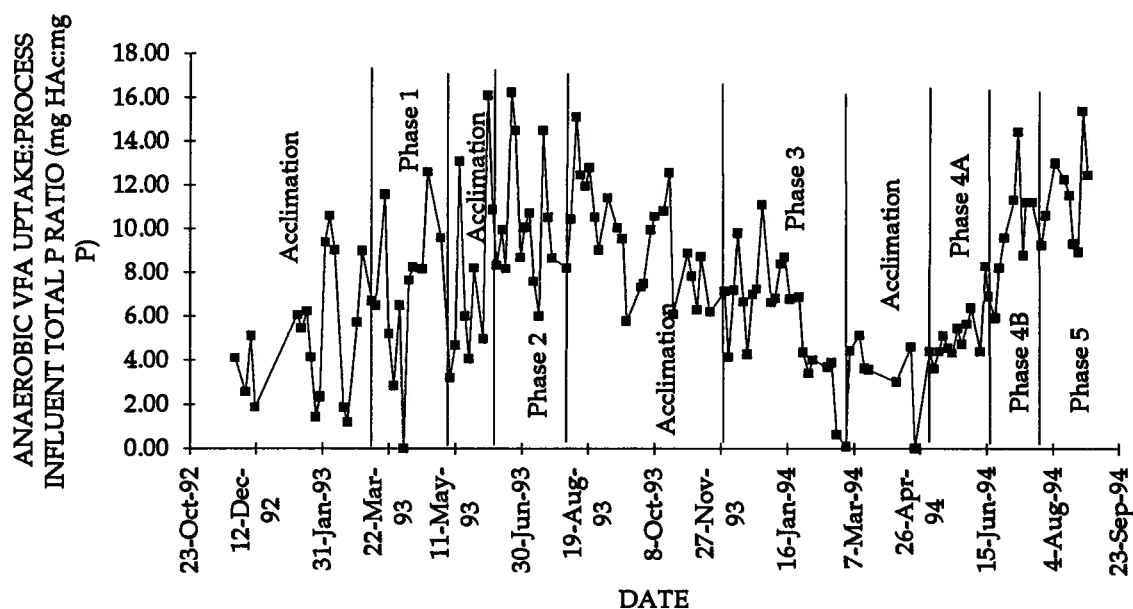
Figure 14 - A Side Anaerobic Reactor VFA Uptake to Process Influent Total
Phosphorus Ratio
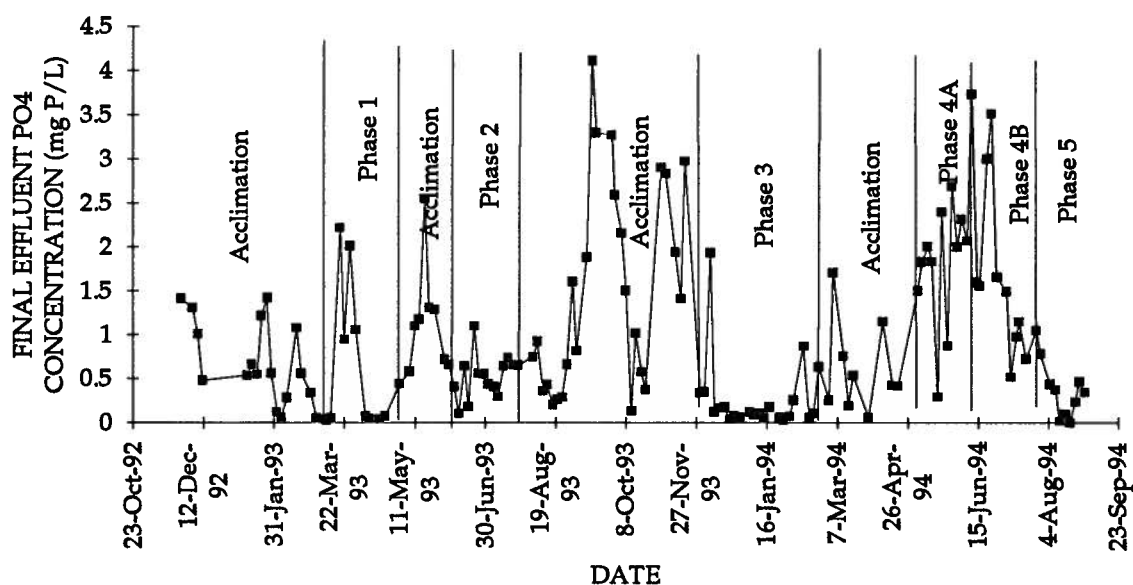
Figure 15 - A Side Process Effluent Orthophosphate Concentration

The pattern of the A Side anaerobic VFA uptake:influent P ratio (Figure 14) is similar
to that shown earlier for the ratio of process influent VFA concentration to process
influent total phosphorus concentration (Figure 11). However, as shown by the results
of pilot plant monitoring during Phases 3 and 4a and the associated batch tests, the

mass of anaerobic VFA uptake during that period was limited by the anaerobic actual HRT, and not by the amount of available VFA or by the VFA uptake capacity (polyP reserves) of the process biomass. Therefore, in this case, the ratio of VFA uptake:influent P is more relevant to effluent phosphorus concentration than the ratio calculated using process influent VFA concentration.

As shown in Figure 14, the ratio of anaerobic VFA uptake to process influent total phosphorus concentration (VFA uptake:influent P ratio) on the A Side was usually in the range 6.5:1 to 12:1 (mg HAc:mg P) during the initial 7 weeks of Phase 3 (November 30, 1993, to January 25, 1994). During the same period, the A Side plant effluent orthophosphate concentration was generally less than 0.2 mg P/L (Figure 15). Near the middle of January, 1994, the VFA uptake:influent P ratio decreased sharply to approximately 4:1 (Figure 14); immediately thereafter, effluent orthophosphate concentration became erratic (Figure 15). Note that the A Side anaerobic actual HRT was 8 minutes throughout Phases 3 and 4a, including the interim acclimation period. During the acclimation period following Phase 3, the VFA uptake:influent P ratio was generally less than 5:1, and plant effluent orthophosphate concentration continued to be erratic, ranging as high as 1.6 mg P/L. During Phase 4a, the VFA uptake:influent P ratio began a steady increase from less than 4:1 to approximately 6:1; during this period, plant effluent orthophosphate concentrations were typically greater than 1.5 mg P/L. At the beginning of Phase 4b, the anaerobic actual HRT was increased to 25 minutes; the VFA uptake:influent P ratio immediately increased into the range 8:1 to 15:1, and plant effluent orthophosphate concentration began a steady decrease. The data in Figures 14 and 15 show that plant effluent orthophosphate concentrations were low when the VFA uptake:influent P ratio was greater than 6:1, and that phosphorus removal deteriorated as the ratio fell below 6:1. The pattern on the B Side was similar to that described above for the A Side.

From the above discussion it is apparent that the degree of phosphorus removal in the pilot plant was limited by the VFA uptake:influent P ratio. During the first 6 weeks of Phase 3, an 8 minute actual anaerobic HRT was adequate to maintain the VFA uptake:influent P ratio at greater than 6:1, and effluent phosphorus concentrations were low, indicating that process phosphorus removal was not limited by anaerobic HRT during that period. An explanation of why the VFA uptake:influent P ratio declined sharply near the middle of January, 1994 (see Figure 14), is given below.

According to Comeau (1989) the rate of VFA uptake in the anaerobic phase should be first order with respect to the amount of polyP reserves in the biomass. As described in Chapter 5, the percent phosphorus by dry weight in the process MLSS provides an estimate of the degree of polyP storage by the process biomass. The percent phosphorus in the A Side aeration basin MLSS is summarized in Figure 16. Note that the phosphorus content of the MLSS decreased sharply in the middle of January, 1994. On January 14, the pipe which carried the complete-mix fermenter overflow to the primary clarifier to mix with the raw plant influent on the A Side became plugged, causing the fermenter to overflow directly into the anaerobic reactor. Nearly the entire solids inventory of the fermenter was carried into the anaerobic reactor, to mix with the process MLSS; the solids inventory in the fermenter at the time of washout was enough to increase the process solids inventory by approximately 35%. Therefore, the sharp decrease in MLSS phosphorus content can be attributed at least partly to dilution of the phosphorus accumulating process biomass with crude (non-phosphorus accumulating) primary solids. Further, the shock load of the fermenter mixed liquor , which contained actively fermenting crude solids and had a bulk solution VFA concentration of approximately 70 mg HAc/L, likely induced an excess of phosphorus release in the system. Phosphorus release under anoxic conditions in response to excess additions of acetic acid has been documented in BNR systems (e.g., Mostert et al., 1988). If the phosphorus release in the pilot plant induced by the shock

load from the fermenter was so great that there was insufficient time for uptake and storage of the released phosphorus before the end of the aerated phase, the polyP content of the process MLSS would further decline.
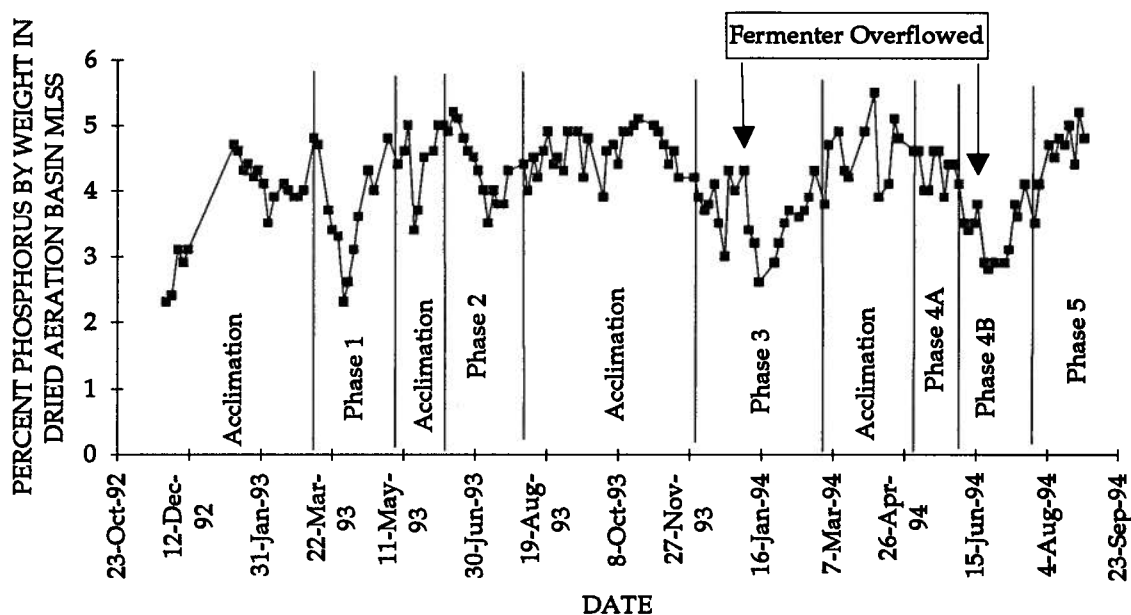
Figure 16 - Percent Phosphorus by Dry Weight in A Side Aeration Basin MLSS

In any case, the sharp drop in process VFA uptake:influent P ratio observed in the middle of January, 1994 (Figure 14), can be attributed to the shock load from the fermenter overflow. The plugging problem with resulting fermenter overflow on the A Side happened on several occasions during the last two weeks of January, 1994. At that point, the primary clarifier was drained, the piping was re-configured, and the operating MLSS concentration of the fermenter was reduced from approximately 4,500 mg/L to approximately 3,000 mg/L. The problem did not occur again until June 20, 1994. At that time, the phosphorus content of the process MLSS again dropped sharply (Figure 16), but in this case there was no corresponding sharp decrease in the VFA uptake to influent P ratio (Figure 14). Note that the anaerobic HRT had been increased to 25 minutes, and the fermenter had a lower solids inventory and a lower

bulk solution VFA concentration (typically 50 mg HAc/L) at the time of the June fermenter washout. The lower shock load in June, combined with the greater anaerobic HRT, apparently had a lesser effect on the process biomass.

During the acclimation period following Phase 3, effluent orthophosphate concentrations on the A Side were erratic (Figure 15). As described in Chapter 5, the B Side anaerobic actual HRT was reduced to 8 minutes at the end of Phase 3, and effluent orthophosphate concentrations immediately became erratic on the B Side as well (see Figure 3b). Apparently, the anaerobic actual HRT of 8 minutes, which was adequate for good phosphorus removal on the A Side during Phase 3, was inadequate during the following acclimation period. As shown in Figure 14, the VFA uptake: influent P ratio on the A Side continued to be less than 5:1 during the acclimation period following Phase 3. The two possible causes of the drop in the VFA uptake: influent P ratio are an increase in process phosphorus loading, and a decrease in anaerobic VFA uptake.

The A Side process influent total phosphorus concentration is summarized in Figure 17. Immediately following Phase 3, there was a large increase in process influent phosphorus concentration from values in the range 2.5-5.0 mg P/L to values greater than 5.0 mg P/L; therefore, the low VFA uptake: influent P ratio during the acclimation period following Phase 3 can be attributed at least partly to an increase in process phosphorus loading.

The mass of VFA taken up in the A Side anaerobic reactor is summarized in Figure 18. As shown, anaerobic VFA uptake on the A Side during the acclimation period following Phase 3 was generally 20 mg HAc/L or less, compared values typically greater than 25 mg HAc/L during Phase 3 (prior to the fermenter overflow on January 14, 1994). Therefore, the relatively low VFA uptake:influent P ratio during the acclimation period following Phase 3 can be attributed to a combination of

increased process phosphorus loading and decreased anaerobic VFA uptake.
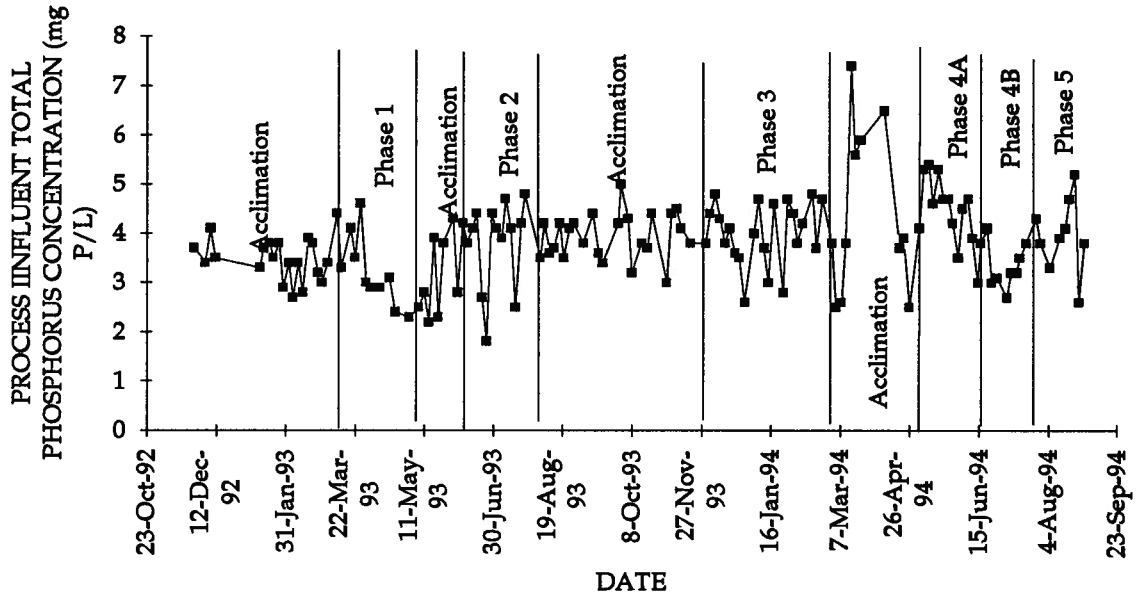
Figure 17: A Side Process Influent Total Phosphorus Concentration
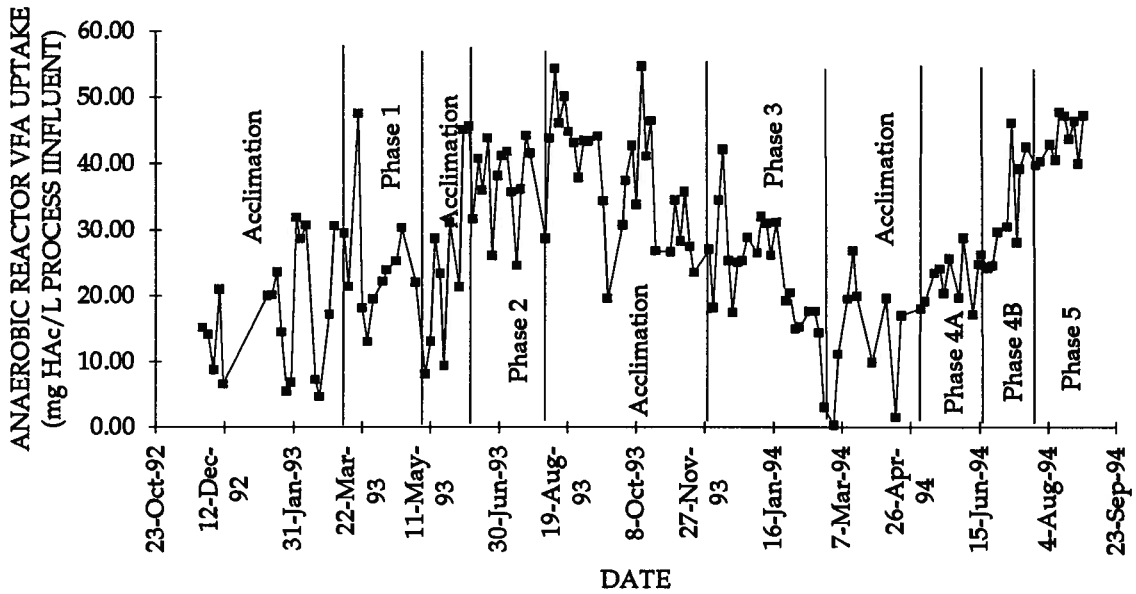
Figure 18: A Side Anaerobic Reactor VFA Uptake (mg HAc/L)

The low VFA uptake during the acclimation period following Phase 3 has two possible explanations. One possibility is a decrease in reactor temperature, which might lead to a slower biochemical reaction rate. However, according to Randall et al.

(1992), biological phosphorus removal processes are relatively insensitive to temperature changes, compared to other biological processes. Further, the temperature of the mixed liquor in the A Side anaerobic reactor was at its lowest during Phase 3 (11-16 °C), and began a steady increase during the following acclimation period. Temperature is therefore not considered to be a factor in the low VFA uptake results.

A second possibility is that a decline in process influent VFA concentration led to a corresponding decrease in anaerobic VFA uptake rate. According to Wentzel et al (1987), the rate of acetate uptake by phosphorus accumulating bacteria in the anaerobic phase is zero order with respect to acetate, and many researchers have demonstrated in bench-scale batch test comparisons that the rate of acetate uptake by the process biomass at a given moment in time is not affected by the concentration of acetate added to the batch reactor (e.g., Comeau, 1989). However, Comeau (1989) demonstrated in bench-scale sequencing batch reactor systems that a system which consistently receives a greater acetate loading to the anaerobic phase develops a higher anaerobic acetate uptake rate; although the anaerobic VFA uptake rate is apparently not affected by short term (e.g., hourly or daily) fluctuations in VFA concentration, the uptake rate may be directly affected by consistently higher or lower VFA concentration over the longer term (e.g., weekly or monthly). The findings of Comeau show that the anaerobic VFA uptake rate increases as the system biomass becomes acclimated to a higher anaerobic VFA concentration.

The concentration of VFA in the A Side process influent is shown in Figure 19. Process influent VFA concentration was generally in the range 30-50 mg/L (as acetic acid) during the period from May of 1993 to mid-January of 1994, and then dropped into the range 15-35 mg/L (Figure 19). The decrease in process influent VFA concentration was caused by the decrease in fermenter steady-state MLSS

concentration instituted following the overflow problems in mid-January, 1994. The data shown in Figures 18 and 19 imply that the lower anaerobic VFA uptake observed during the acclimation period following Phase 3 was due to a lower VFA uptake rate, which in turn was possibly caused by a lower VFA concentration at that time. It was not simply that insufficient VFA was present, because there was still VFA left at the end of the anaerobic phase.

During Phase 4a, process influent VFA concentrations increased slightly (Figure 19), the mass of VFA taken up in the anaerobic reactor became much more consistent (Figure 18), and process phosphorus loading began to decrease (Figure 17). Accordingly, the VFA uptake:influent P ratio began a steady increase during that period, but was still well below 6:1 most of the time (Figure 14), and effluent orthophosphate concentrations were typically well over 1 mg P/L.
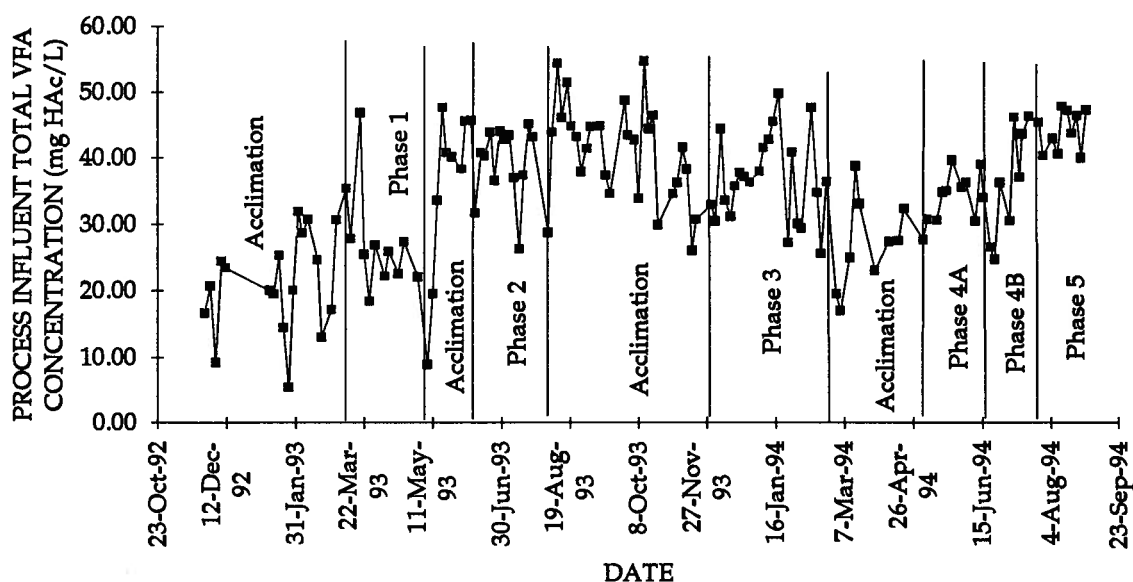
Figure 19: A Side Process Influent Total VFA Concentration

At the end of Phase 4a, the anaerobic actual HRT was increased from 8 minutes to 25 minutes, the VFA uptake:influent P ratio increased sharply to values greater than 8:1 (Figure 14), and effluent orthophosphate concentration began a steady decrease

(Figure 15). During Phase 5, the VFA uptake:influent P ratio was consistently greater than 8:1, and effluent orthophosphate concentrations declined to less than 0.5 mg P/L. The phosphorus content of the aeration basin MLSS increased steadily throughout Phases 4b and 5, except for the decrease induced by the fermenter overflow in June, 1994 (Figure 16).

On the B Side process, effluent phosphorus concentration declined during Phase 4b, in a similar manner to the A Side. However, immediately following the reduction in MLSS concentration on the B Side at the outset of Phase 5, effluent phosphorus concentration on that side began to increase (see Figure 3). The results of pilot plant monitoring showed that the mass of anaerobic VFA uptake and the phosphorus content of the aeration basin MLSS were significantly higher on the A Side during Phase 5 (Tables 12 and 13, respectively). The mean VFA concentration in the anaerobic reactor on the B Side was 4.51 mg HAc/L, compared to only 0.19 mg HAc/L on the A Side during Phase 5. The anaerobic VFA uptake:influent P ratio on the B Side increased immediately into the range 8:1 to 15:1 at the outset of Phase 4b, in a similar manner to the A Side. However, the ratio on the unacclimated B Side declined immediately following the reduction in MLSS concentration, and was less than 6:1 on 4 out of 8 sampling scans during Phase 5. Therefore, it is likely that the 25 minute anaerobic actual HRT was not adequate for sufficient VFA uptake at the lower operating MLSS concentration on the B Side. That is, phosphorus removal in the pilot plant was not limited by a 25 minute anaerobic actual HRT at an operating aeration basin MLSS concentration of approximately 3,000 mg/L, but was limited by a 25 minute anaerobic HRT at an operating MLSS concentration of 2,000 mg/L. T As noted earlier, there was no acclimation period between Phases 4 and 5, and Phase 5 was only one month long. If the B Side had been allowed a longer acclimation period, the percentage of phosphorus accumulating bacteria might have increased, with a resulting increase in the VFA uptake rate. The result would be an increase in the

percent phosphorus in the process biosolids, which in turn would increase the amount of phosphorus wasted, and lead to lower plant effluent phosphorus concentrations. he effects of operating MLSS concentration on process performance are discussed in more detail later in this thesis.

The data discussed above show that the effect of the anaerobic actual HRT on phosphorus removal depends in part on the process influent VFA concentration, and can also depend on the anaerobic reactor MLSS concentration. At high influent VFA concentrations (35-50 mg HAc/L), an anaerobic actual HRT of 8 minutes was adequate for good phosphorus removal. However, at lower influent VFA concentrations (20-40 mg HAc/L), the 8 minute HRT was not long enough for adequate VFA uptake, due to a resulting decrease in the anaerobic VFA uptake rate. The 25 minute anaerobic HRT was more than adequate at an anaerobic reactor mean MLSS concentration of 2121 mg/L, but was inadequate at an anaerobic reactor mean MLSS concentration of 1464 mg/L. The changes in process performance discussed above indicate that increased opportunities for anaerobic carbon storage led to increases in the mass of anaerobic VFA uptake and the phosphorus content of the solids, and consequent decreases in effluent phosphorus concentration.

6.3 DENITRIFICATION AND PHOSPHORUS UPTAKE IN THE ANOXIC PHASE

As described in Chapter 5, denitrification in the anoxic reactor was usually associated with phosphorus uptake, indicating that the bacteria responsible for polyP storage were able to use nitrate as an electron acceptor in the absence of molecular oxygen, for the oxidation of PHA reserves and synthesis of polyP. The batch test results showed that denitrification continued at a linear rate until all nitrates were removed from solution, and in most cases, the associated phosphorus uptake was also linear. The batch tests also showed that allowing significant quantities of VFA to penetrate to

the anoxic phase decreased the net anoxic phosphorus uptake (see Figures 4-10). Further, as described in the Literature Review, the addition of VFA (e.g., acetate and propionate) to the anoxic phase of biological treatment systems has been shown to enhance the denitrification rate. The effects of VFA carryover from the anaerobic to the anoxic phase on denitrification and phosphorus uptake in the pilot plant are discussed in the following sections.

6.3.1 The Effects of VFA Carryover from the Anaerobic Phase on the Anoxic Denitrification Rate

A parallel comparison of the effects of a higher concentration of VFA in the process mixed liquor entering the anoxic reactor is possible for Phase 3, when the A Side anaerobic actual HRT of 8 minutes resulted in a significantly higher mean concentration of VFA in the process stream entering the A Side anoxic reactor compared to the B Side, which had an anaerobic actual HRT of 25 minutes (i.e., 5.30 mg HAc/L on the A Side compared to 1.76 mg HAc/L on the B Side - see Table 12). A total of four batch tests were carried out during Phase 3. The concentration of VFA remaining at the end of the anaerobic phase, the average anoxic denitrification rates, and the average specific anoxic denitrification rates for those batch tests are compared using the t test in Table 18. In this case, the one-tailed t test at the 0.05 level of significance was used, since the results of others showed that a a higher VFA concentration entering the anoxic reactor could be expected to lead to a faster denitrification rate.

As shown in Table 18, the mean concentration of VFA at the end of the anaerobic phase of the batch tests was significantly greater on the A Side (6.79 mg HAc/L) than on the B Side (0.83 mg HAc/L). The mean denitrification rate was higher on the A Side (4.91 mg N/L/hr) than on the B Side (3.92 mg N/L/hr), but the difference was not significant according to the t test. The specific denitrification rate was also higher

on the A Side (1.42 mg N/hr/g MLSS) than on the B Side (1.31 mg N/hr g MLSS), but again the difference was not statistically significant. Note that the specific denitrification rate was higher on the A Side in only two out of four cases.

Table 18 - Effects of VFA Concentration on Denitrification Rate - Phase 3

Date of Test	VFA Concentration at End of Anaerobic Phase (mg HAc/L)		Average Anoxic Denitrification Rate (mg N/L/hr)		Average Specific Anoxic Denitrification Rate (mg N/hr/g MLSS)	
	A Side	B Side	A Side	B Side	A Side	B Side
Dec 2/94	4.72	1.70	5.80	4.69	1.82	1.50
Dec 16/94	3.67	0.00	4.17	3.73	1.17	1.23
Jan 10/94	12.20	0.81	6.28	3.71	1.45	1.15
Feb 7/94	6.58	0.79	3.39	3.58	1.25	1.35
Mean	6.79	0.83	4.91	3.92	1.42	1.31
calculated t	3.14		1.65		1.02	
critical t (one-tailed test)	2.35		2.35		2.35	
Significant Difference?	Yes		No		No	

The data in Table 18 show that allowing significant quantities of VFA to reach the anoxic reactor did not significantly enhance the denitrification rate, and are therefore not consistent with the findings of others as described in the Literature Review. Possibly, a higher specific denitrification rate on the A Side would have been detected if a larger number of batch tests had been performed. On the other hand, most of the batch tests conducted throughout the study showed that denitrification was accompanied by phosphorus uptake, indicating that at least some nitrate reduction was carried out by phosphorus accumulating bacteria. Since those bacteria use stored PHA as a carbon source, the denitrification rate of the phosphorus accumulators

might not be affected by the nature and concentration of bulk solution organic carbon. The relative proportion of phosphorus accumulating denitrifiers to non-phosphorus accumulating denitrifiers in the pilot plant might then have determined the extent to which the VFA concentration affected the overall denitrification rate.

6.3.2 The Effects of VFA Carryover from the Anaerobic Phase on Anoxic Phosphorus Removal

Phosphorus release in biological phosphorus removal systems under anoxic conditions in response to additions of acetate has been documented (e.g., Mostert et al., 1988 and Comeau et al., 1987). Mostert et al. (1988) speculated that simultaneous release and uptake of phosphorus occurred in the presence of acetate in the anoxic phase, and that phosphorus accumulating bacteria competed with denitrifiers for the available organic substrate. Comeau et al. (1987) suggested that the microbial population in a BNR plant contains some species of phosphorus accumulating bacteria which are capable of denitrification, and some which are not. In that case, both phosphorus accumulating and non-phosphorus accumulating denitrifiers would use nitrate as an electron acceptor, with phosphorus uptake and PHA consumption being performed by the phosphorus accumulating denitrifiers. Non-denitrifying (obligate aerobic) phosphorus accumulating bacteria would behave as though under anaerobic conditions; that is, they would continue to take up and store VFA, with associated phosphorus release. The net result of the anoxic activity would then depend on the relative numbers of obligate aerobic and denitrifying phosphorus accumulating bacteria.

In batch tests where the concentration of VFA at the end of the anaerobic phase was approximately 2 mg HAc/L or less, anoxic phosphorus release was not observed. In batch tests where the concentration of VFA at the end of the anaerobic phase was as high as 3-4 mg HAc/L, simultaneous denitrification, net phosphorus release, and

removal of VFA from solution were often observed during the first five minutes of the anoxic phase; thereafter, denitrification was associated with phosphorus uptake in a similar manner to the other batch tests (see Figures 6a, 7a, 7b, 9c, and 10a). The batch test results suggest that allowing high concentrations of VFA to penetrate to the anoxic reactor reduced the net anoxic phosphorus uptake. The results of pilot plant monitoring during Phase 3 support the batch test data. Mass balances on the anoxic reactor showed that, during Phase 3, the mean VFA concentration in the anaerobic reactor was 5.3 mg HAc/L on the A Side, compared to only 1.8 mg HAc/L on the B Side (Table 12); corresponding mean orthophosphate uptake in the anoxic reactor on the A and B Sides was 3.6 mg P/L and 14.2 mg P/L, respectively. Similarly, during Phase 4, a higher mean concentration of VFA in the anaerobic reactor on the B Side (3.5 mg HAc/L compared to 2.8 mg HAc/L on the A Side - see Table 12) was associated with a lower mean anoxic phosphorus uptake (7.9 mg P/L on the B Side compared to 10.5 mg P/L on the A Side - see Table 14). The data show that allowing high concentrations of VFA to enter the anoxic phase during Phases 3 and 4 reduced the net anoxic phosphorus uptake.

The data discussed above (i.e., noticeable phosphorus release in the anoxic zone only at VFA concentrations greater than 3 mg HAc/L) appear contradictory to the findings of others who have shown that VFA uptake with associated phosphorus release is zero order with respect to VFA concentration. However, a possible explanation for the pilot plant data might be that persistent high VFA concentrations entering the anoxic zone (A Side during Phase 3 and both sides during Phase 4a) provided non-denitrifying phosphorus accumulators with an additional opportunity for carbon storage. Over time, the percentage of non-denitrifying phosphorus accumulators in the system might then increase, as the biomass became acclimated to high anoxic VFA concentrations. The result would be an increase in the amount of phosphorus released in the anoxic zone in response to VFA additions.

On the other hand, during Phase 5, when the pilot plant anaerobic mean VFA concentration was only 0.2 mg HAc/L on the A Side compared to 4.5 mg HAc/L on the B Side (Table 12), anoxic phosphorus uptake was similar on both the A and B Sides (8.3 mg P/L and 7.4 mg P/L, respectively - Table 14). The data show that allowing significant amounts of VFA to enter the anoxic reactor during Phase 5 did not induce significant anoxic phosphorus release (with a consequent reduction in anoxic phosphorus uptake). Possibly, the higher sludge wasting rate on the B Side (mean=210 L/day) compared to the A Side (mean=110 L/day) during Phase 5 resulted in a change in the bacterial population of the suspended biomass. If non-denitrifying phosphorus accumulators tended to be washed out of the system at the lower sludge age, uptake and storage of VFA with associated phosphorus release would not be observed in the anoxic reactor; this might explain the lack of anoxic phosphorus release on the B Side, despite high initial concentrations of VFA. The lower phosphorus content of the process biomass on the B Side (3.4%) compared to the A Side (4.6% - see Table 13) during Phase 5 implies a lesser degree of polyP storage, whether due to a lower percentage of bacteria capable of phosphorus accumulation, or to a lesser degree of polyP storage by individual bacteria. In either case, the lower overall bacterial polyP reserves on the B Side during Phase 5 might have slowed the overall rate of anoxic VFA uptake-phosphorus release, allowing non-phosphorus accumulating denitrifiers to use most of the available VFA; the net result would be a relatively low degree of anoxic phosphorus release.

In any case, the data show that allowing significant quantities of VFA to enter the anoxic reactor when the pilot plant was operated at an aeration basin MLSS concentration of approximately 3,000 mg/L induced net anoxic phosphorus release during the first few minutes of denitrification, and thereby reduced the overall anoxic phosphorus uptake. If the anoxic phosphorus release was the result of obligate aerobic phosphorus accumulating bacteria taking up and storing the available VFA,

the released phosphorus should have been taken up and stored as polyP in the aerobic reactors. However, since at least some of the VFA entering the anoxic reactor was probably used by non-phosphorus accumulating denitrifiers, it would be preferable to maximize VFA uptake in the anaerobic phase, and minimize VFA carryover to the anoxic phase, if phosphorus uptake in the anoxic and anaerobic zones is to be maximized.

As described above, the batch test results showed that allowing significant concentrations of VFA to enter the anoxic phase resulted in phosphorus release during the initial stages of denitrification. Therefore, the addition of VFA to the anoxic phase (with consequent phosphorus release) could mask the true ratio of phosphorus taken up to nitrate reduced. For those batch tests in which the concentration of VFA at the end of the anaerobic phase was less than 1 mg HAc/L (including tests conducted during the interim acclimation periods), the mean molar ratio of phosphorus uptake to NO_x reduction in the anoxic phase was 0.55 on the A Side (standard deviation=0.19) and 0.60 on the B Side (standard deviation=0.20). No detailed theoretical or empirical discussions of nitrate demand during anoxic phosphorus uptake were found in the literature for comparative purposes. However, the results of this study show that the anoxic reactor can play a significant role in biological phosphorus removal, especially if little VFA escapes from the anaerobic zone to the anoxic zone. Mass balance calculations on all of the pilot plant data collected during Phases 1-5 and the interim acclimation periods showed that an average of 45% of the total process phosphorus uptake occurred in the anoxic reactor. Further study is needed to better define bacterial phosphorus uptake and storage under anoxic conditions in BNR systems, particularly the relative amounts of PHA consumption to phosphorus uptake in the anoxic and aerobic phases; also important is the relative amount of anoxic nitrate consumption to phosphorus uptake-storage, so that the role of the anoxic reactor can be defined in light of the economics and effluent

restrictions of full-scale BNR plants.

6.3.3 The Effects of HRT in the Anoxic Phase on Denitrification and Phosphorus Removal

The effects of anoxic HRT were investigated during Phase 1, when an anoxic actual HRT of 65 minutes on the A Side was compared to one of 35 minutes on the B Side. The batch test results showed that, as soon as nitrates were removed from solution in the anoxic phase, phosphorus release generally began (see Figures 4, 5, 6, 7, 8, 9, and 10). As described in the previous section, phosphorus removal associated with denitrification implies that phosphorus accumulating bacteria can degrade PHA reserves using nitrate as an electron acceptor to rebuild polyP reserves. Simultaneous PHA degradation and polyP accumulation under anoxic conditions has been documented (e.g., Comeau et al., 1987). However, preliminary results reported by Comeau et al. (1987) showed that only 2.6 moles of phosphate were accumulated per mole of PHA consumed using nitrate as an electron acceptor, compared to 3.3 moles of phosphate accumulated per mole of PHA consumed using oxygen. Comeau attributed the difference to the lower theoretical energy yield using nitrate as an electron acceptor, compared to oxygen. In a parallel comparison using bench-scale sequencing batch reactor systems, Vlekke et al. (1988) found that the total amount of PHA consumed during phosphorus uptake was 50% greater with nitrate as the electron acceptor than with oxygen. If the results of Comeau et al. (1987) and Vlekke et al. (1988) provide a true indication of comparative amounts of PHA consumption using nitrate and oxygen as the electron acceptor, and if phosphorus removal is the primary objective, it might be advisable to limit denitrification in a BNR plant to that which is required to prevent nitrates from reaching the anaerobic reactor, since further denitrification might lead to the consumption of excess PHA reserves. Ultimately, the optimum degree of denitrification would then depend on the relative amounts of VFA, phosphorus, and nitrogen in the process influent, and on the

treatment priorities of the process.

In any case, the pilot plant results show that the anoxic reactor can be integral to the optimization of a BNR process. The beneficial effect of minimizing the anoxic retention time to that required for complete denitrification is clearly shown by the Phase 1 batch test results (Figure 4). For the A side (45 minute anoxic actual HRT), the phosphorus release in the anoxic zone following the completion of denitrification resulted in a phosphorus concentration of 5.4 mg P/L at the outset of the aerobic phase (Figure 4a). Bacterial phosphorus uptake resulting in a concentration of less than 0.5 mg P/L on the A side required 80 minutes of aeration. On the B Side (25 minute anoxic actual HRT), where considerably less post-denitrification phosphorus release occurred, the orthophosphate concentration at the outset of the aerobic phase was only 2.3 mg P/L, and only 10 minutes of aeration was required to reach an orthophosphate concentration of less than 0.5 mg P/L (Figure 4b). Therefore, it is evident that, to minimize post-denitrification phosphorus release, the actual HRT of the anoxic reactor should be limited to that which is required for complete denitrification. In a full-scale plant where flow and load variations are the norm, optimum nitrate loading of the anoxic reactor might be accomplished through on-line manipulation of nitrate-bearing process internal recycle rate(s) from the final clarifier and/or aerobic reactor to the anoxic reactor.

According to the Phase 1 batch test results described above, phosphorus removal in the pilot plant anoxic reactor should have been greater on the B Side than on the A Side during Phase 1. However, mean anoxic orthophosphate uptake was significantly greater on the A Side (13.0 mg P/L) than on the B Side (11.1 mg P/L) during Phase 1 (see Table 14). This apparent anomaly can be explained by the significantly higher mean mass of NO_x reduction on A Side (8.1 mg N/L), compared to the B Side (6.8 mg N/L); the greater NO_x reduction on the A Side during Phase 1 can be attributed to a

higher NO_x concentration in the return settled biosolids stream (Table 11). Further, the mean anoxic MLSS concentration during Phase 1 was 2580 mg/L on the A Side, compared to 2940 mg/L on the B Side - see Table 6). The combination of a higher NO_x loading and a lower operating MLSS concentration in the A Side anoxic reactor could be expected to result in a longer time required for complete denitrification on the A Side, reducing the time available for post-denitrification phosphorus release, and thereby increasing the net anoxic phosphorus uptake.

6.4 NITRIFICATION AND PHOSPHORUS UPTAKE IN THE AEROBIC PHASE

6.4.1 Fixed growth Nitrification vs. Suspended growth Nitrification

As described in Chapter 5, greater than 85% of the nitrification in the pilot plant occurred in the FGRs; nitrate production in the aeration basin was usually less than 1 mg N/L, compared to 9.6-12.2 mg N/L in FGR Cell 1, and 2.3-6.3 mg N/L in FGR Cell 2 (see Table 11). The ammonia concentration in the mixed liquor leaving FGR Cell 2 was usually less than 1 mg N/L. However, as shown in the batch test results (Figures 4-10), nitrifying organisms were present in suspension in the process mixed liquor as well. Suspended nitrifiers were probably a mixture of fixed growth organisms which sloughed from the FGR media, and organisms which grew in suspension in the mixed liquor. Therefore, at least some of the nitrification observed across each FGR cell must have been accomplished by suspended organisms in the underflow catchbasin and in the process mixed liquor cascading over the media. The relative amounts of nitrification by the fixed and suspended growth components of the pilot plant are estimated below.

The mean suspended growth specific nitrification rates for all steady-state batch test simulations conducted after pulse loading of the FGRs was begun (i.e., from November 30, 1993, through August 30, 1994) are summarized in Table 19. As shown,

the mean suspended growth specific nitrification rate was 1.38 mg N/hr/g MLSS on the A Side, and 1.39 mg N/hr/g MLSS on the B Side. According to Randall et al. (1992), typical specific nitrification rates in BNR systems at operating MLSS concentrations in the range 2,500-3,500 mg/L are 1.8-2.0 mg N/hr/g MLVSS. The pilot plant MLVSS represented an average of 74% (standard deviation=2.2%) of the total MLSS on the A Side, and an average of 72% (standard deviation=3.4%) of the total MLSS on the B Side, during the period November 30, 1993, through August 30, 1994. On the basis of MLVSS, the mean specific suspended growth nitrification rates shown in Table 18 are 1.86 mg N/hr/g MLVSS for the A Side, and 1.93 mg N/hr/g MLVSS for the B Side, similar to those reported by Randall et al. (1992).

Table 19 Specific Suspended Growth Nitrification Rates from Steady-State Batch Test Simulations (November 30/93 to August 30/94)

Specific Suspended Growth NO_x Production Rate (mg N/hr/g MLSS)					
Process	Mean	Std. Dev.	Minimum	Maximum	Sample n
A Side	1.38	0.45	0.84	2.28	20
B Side	1.39	0.40	0.76	2.19	20

As described earlier in Sections 5.1.3 and 5.1.4, the mean ammonia concentration in the influent to the FGRs was at or below the first order threshold of 3-5 mg N/L, and the FGRs therefore operated in the regime where fixed growth nitrification rates are first order with respect to ammonia concentration. Note that the mean ammonia concentration in the process mixed liquor entering FGR Cell 2 was always less than 1 mg N/L, while the ammonia concentration entering FGR Cell 1 was in the range 1.7-3.0 mg N/L (Table 10). The mass balances on the FGR Cells showed that the mean NO_x production in FGR Cell 2 was much lower than that in FGR Cell 1 for both the A

and B Sides (i.e., 9.6-12.2 mg N/L in Cell 1 compared to 2.3-6.3 mg N/L in Cell 2 - see Table 11), confirming that the nitrification rate in the FGRs decayed with decreasing ammonia concentration.

On the other hand, the batch test results showed that the suspended growth nitrification rate in the pilot plant was essentially zero order with respect to ammonia concentration (see Figures 4-10). Therefore, it is likely that the suspended growth in the process mixed liquor as it passed through the FGR catchbasins and cascaded over the media accounted for a greater percentage of the total mass of ammonia nitrified in FGR Cell 2 than in Cell 1.

The nominal HRT in each FGR underflow catchbasin at steady-state flow rates was 40-45 minutes. The amount of nitrification accomplished by the suspended organisms in the mixed liquor in each FGR Cell catchbasin was estimated using the mean specific suspended growth nitrification rate shown in Table 19 (i.e., 1.4 mg N/hr/g MLSS), the nominal hydraulic retention time in each FGR catchbasin, the estimated nominal hydraulic retention time on the FGR media (assumed to be five minutes - see Table 15), and the MLSS concentration in the aeration basin. The estimated degree of suspended growth nitrification in each FGR Cell is compared to the total nitrate production from reactor mass balances for that cell for the period November 30, 1993, through August 30, 1994, in Table 20. As shown, the fixed growth accounted for an estimated 65% and 68% of the total nitrification across FGR Cell 1 on the A and B Sides, respectively. Again assuming a zero-order suspended growth specific nitrification rate of 1.4 mg N/hr/g MLSS - from Table 19), the fixed growth accounted for an estimated 24% and 40% of the total nitrification across FGR Cell 2 on the A and B Sides, respectively (Table 20).

Table 20 Estimated Fixed Growth Nitrification vs. Suspended growth Nitrification
(November 30/93 to August 30/94)

Nitrate Production (mg N/L Plant Influent)				
Process Reactor	1 Suspended Growth[1]	2 Total FGR Cell Including Catchbasin[2]	3 Fixed Growth[3]	4 Percentage of Nitrification by Fixed Growth[4]
	Mean (Std. Dev.)	Mean (Std. Dev.)		
A Side Cell 1	3.5 (0.7)	10.8 (2.0)	7.3	68%
B Side Cell 1	3.3 (0.8)	9.3 (1.8)	6.0	65%
A Side Cell 2	3.5 (0.7)	4.6 (1.6)	1.1	24%
B Side Cell 2	3.3 (0.8)	5.5 (1.4)	2.2	40%

[1] Based on the nominal HRT of the process mixed liquor in the FGR catchbasin and through the FGR tower, a specific suspended growth nitrification rate of 1.4 mg N/hr/g MLSS, and the aeration basin MLSS concentration
[2] Based on mass balance of individual FGR cell - see Table 11
[3] Estimated by subtracting value in column 1 from value in column 2
[4] Value in column 3 as a percentage of value in column 2

It is apparent that the pilot plant had an excess of nitrification capacity. As described above, the suspended growth in the pilot plant exhibited nitrification rates similar to those typical of strictly activated sludge-type BNR plants. However, little nitrification occurred in the aeration basin, since ammonia was generally nitrified in the FGRs to less than 1 mg N/L. Further, the ammonia concentration in the mixed liquor irrigated over the FGRs was relatively low (due to relatively low process influent ammonia concentrations, typically in the range 10-12 mg N/L), resulting in a fixed growth nitrification rate which decayed with decreasing ammonia concentration. To optimize

the design of the FGR-SGR process for nitrification, the FGRs should be sized such that the nitrification capacity of the aeration basin is utilized (i.e., the degree of nitrification in the FGRs should be limited to that which allows residual ammonia in the process mixed liquor leaving the FGRs to be removed to the desired level in the aeration basin). Further, to maximize the fixed growth nitrification rate, the design should be such that the mixed liquor leaving the FGR media has an ammonia concentration at or above the first order threshold of 3-5 mg N/L.

6.4.2. Phosphorus Uptake in the FGRs

Phosphorus accumulating bacteria were not expected to grow in the biofilm attached to the FGR media, since fixed growth bacteria would not have the opportunity to take up and store VFA under anaerobic conditions. From the standpoint of phosphorus removal, the FGRs primary function was cascade aeration of the process mixed liquor. Therefore, phosphorus uptake in the FGRs would be due to normal metabolic requirements, and phosphorus uptake and storage by suspended bacteria in the process liquid cascading over the media. Phosphorus uptake and storage by suspended bacteria would also be expected to occur under aerobic conditions by suspended organisms in the FGR catchbasins and effluent sump. The phosphorus removal across each FGR cell in the pilot plant is compared to the batch test results below.

The actual HRT in each FGR underflow catchbasin at steady-state flow rates was 20-23 minutes. As described in Chapter 5, the actual HRT of the process liquid on the FGR media was estimated at approximately 1 minute (Table 15). The mean specific orthophosphate uptake rate across FGR Cell 1 in the pilot plant is compared to the mean specific orthophosphate uptake rate during the first 20 minutes of all batch tests conducted after the initiation of pulse hydraulic loading of the FGRs in Table 21. Similarly, the mean specific orthophosphate uptake rate across FGR Cell 2 in the pilot

plant is compared to the mean specific orthophosphate uptake rate during the second 20 minutes of all batch tests during the same period.

Table 21 Comparison of Specific Phosphorus Uptake Rate in the Fixed Growth Reactors vs. Batch Tests (November 30/93 to August 30/94)

Specific Orthophosphate Uptake Rate (mg P/hr/g MLSS)						
Process Train	Reactor	Mean	Standard Deviation	Range		Sample n
				Low	High	
A Side	FGR Cell 1[1]	3.27	2.00	0	9.50	63
	Batch (0-20 min)[2]	4.36	2.46	1.43	9.69	20
	FGR Cell 2[1]	1.44	1.30	0	4.90	63
	Batch (20-40 min)[3]	1.97	1.19	0.69	5.40	20
B Side	FGR Cell 1[1]	3.36	1.92	0	8.87	61
	Batch (0-20 min)[2]	3.44	1.19	0.78	7.46	20
	FGR Cell 2[1]	1.90	1.57	0	4.00	62
	Batch (20-40 min)[3]	1.69	1.14	0	3.75	20

[1] Based on the actual HRT of the process mixed liquor in the FGR catchbasin (i.e., approximately 20 minutes), where the actual HRT of the mixed liquor on the FGR media is assumed to be 1 minute and is not included- see Table 15)

[2] Based on the initial 20 minutes of the aerated phase in all batch tests conducted between November 30, 1993, and August 30, 1994

[3] Based on the second 20 minutes of the aerated phase in all batch tests conducted between November 30, 1993, and August 30, 1994

As shown in Table 21, the specific orthophosphate uptake rates observed in both the batch tests and in the FGRs ranged from less than 1 mg P/hr/g MLSS to greater than 9 mg P/hr/g MLSS. The mean uptake rates across the A Side FGR Cells 1 and 2 were slightly lower than the mean uptake rates during the first 20 minutes of the A Side

batch tests (i.e., 3.3 mg P/hr/g MLSS across FGR Cell 1, compared to 4.4 mg P/hr/g MLSS during the first 20 minutes of the batch tests, and 1.4 mg P/hr/g MLSS across FGR Cell 2, compared to 2.30 mg P/hr/g MLSS during the second 20 minutes of the batch tests). On the B Side, the FGR uptake rates were similar to those of the batch tests (i.e., 3.4 mg P/hr/g MLSS across FGR Cell 1, compared to 3.4 mg P/hr/g MLSS during the first 20 minutes of the batch tests, and 1.9 mg P/hr/g MLSS across FGR Cell 2, compared to 1.7 mg P/hr/g MLSS during the second 20 minutes of the batch tests). Similar batch tests conducted during the full-scale study at Salmon Arm resulted in a mean specific aerobic orthophosphate uptake rate of 4.1 mg P/hr/g MLSS, similar to the results reported above for the initial 10 minutes of aeration in the pilot plant batch tests. The exponential decay in aerobic phosphorus uptake rate typical of the pilot plant batch tests was not observed in the Salmon Arm batch tests; however, the specific phosphorus uptake rate across the FGR and underflow catchbasin at the Salmon Arm plant (7.7 mg P/hr/g MLSS - based on the HRT of the catchbasin only) was higher than that in the downstream aeration basin (1.2 mg P/hr/g MLSS - Gibb, 1990).

Note that the mean specific orthophosphate uptake rates across FGR Cells 1 and 2 were similar between the A and B Sides, and both were similar to the batch test results on the B Side (Table 21). Although the mean specific orthophosphate uptake rates during the first two 20 minute increments of the batch tests were slightly higher on the A Side than on the B Side, a t test comparison between the two showed that the difference in means was not significant.

6.4.3 The Effect of FGR Recycle Rate on Phosphorus Removal

As described in Chapter 5, manipulation of the FGR recycle rate had a profound effect on process phosphorus removal. Solids accumulation on the FGR media observed during the period between Phases 2 and 3 led to erratic phosphorus removal; plant

effluent orthophosphate concentrations during that period ranged from approximately 0.1 mg P/L to greater than 4 g P/L (see Figures 3a, 3b, and 15). The mean percent phosphorus by weight in the process MLSS during the period between Phases 1 and 2 was 4.6% on both the A and B Sides, as high or higher than during other periods where phosphorus removal was consistently good (e.g., see Phases 1-3 in Table 13). Therefore, the deterioration in phosphorus removal during the interim period between Phases 2 and 3 could not have been due a change in the phosphorus storage capacity of the process biomass.

Except for the acclimation period between Phases 2 and 3, the average daily volume of mixed liquor wasted from the aeration basin was typically in the range 100-200 L. As described in Chapter 5, sludge wasting was reduced and periodically discontinued altogether during the period between Phases 2 and 3, in an effort to maintain the aeration basin MLSS concentration at approximately 3,000 mg/L. The resulting average daily volume of mixed liquor wasted from the aeration basin during the interim period between Phases 2 and 3 was 31 L on the A Side, and 66 L on the B Side. Therefore, the relatively poor phosphorus removal during that period can be attributed to lower masses of phosphorus wasted from the process on a daily basis.

Immediately following the beginning of pulse hydraulic loading to the FGRs, process MLSS concentrations were much easier to control, daily sludge wasting volumes increased and became more consistent, and phosphorus removal improved dramatically. It is therefore apparent that control of solids accumulation in the FGRs by periodic hydraulic scouring of the media is an important operational parameter for the FGR-SGR process. At the full-scale FGR-SGR biological phosphorus removal plant at Salmon Arm, where the hydraulic loading rate is similar to that of the pilot plant, excess solids build up on the FGR media has not been observed during steady-state

operation since 1988. However, the FGR media at Salmon Arm is of a different physical configuration than the media used in the pilot plant. Therefore, it appears that media configuration can have a significant effect on the need for pulse loading in the FGR-SGR process.

6.4.4 The Effect of FGR Recycle Rate on Nitrification

Contrary to the results for phosphorus removal described in the previous section, the institution of the pulse loading regime for the FGRs did not cause a noticeable change in nitrification across any of the individual FGR cells. Accumulation of excess solids on the FGR media would be expected to slow the flux of ammonia to the fixed biofilm as it was buried by accumulating solids, thereby reducing the nitrification rate. However, the batch tests showed that the process suspended solids contained a significant population of nitrifiers. Therefore, fixed growth nitrifiers buried as the biofilm increased in thickness due to solids accumulation on the FGR media would continuously be replaced at the water-solids interface by nitrifiers in the process MLSS. In any case, scouring of the FGR media by a daily increase in FGR recycle rate did not significantly enhance or reduce nitrification in the FGRs.

The comparison of the effects of the steady-state FGR irrigation rate carried out during Phase 4 showed that there was no statistically significant difference in FGR ammonia removal at an irrigation rate of 24 L/min (A Side), compared to one of 14 L/min (B Side). However, the mean ammonia removal during Phase 4 was lower across the B Side FGR Cell 1 (5.81 mg N/L), compared to the A Side (7.14 mg N/L - see Table 10). The large standard deviations in ammonia removal across FGR Cell 1 during Phase 4 (approximately equal to the means for both of the A and B Sides) reduced the chances of detecting a significant difference using the t test. The corresponding mean NO_x production across FGR Cell 1 was 10.19 mg N/L across the B Side FGR Cell 1, compared to 11.97 mg N/L on the A Side (Table 11); in this case,

the standard deviations were much lower (20-25% of the means), and the difference was found to be statistically significant, according to the t test.

In any case, as described in Section 6.4.1, the pilot plant had an excess of nitrification capacity because of the low ammonia loading to the FGRs, and process ammonia removal to concentrations less than the detection limit of 0.01 mg N/L were consistent for both sides throughout Phase 4. However, in a process more heavily loaded with ammonia, where detectable concentrations of ammonia were present in the plant effluent, the greater ammonia removal associated with the higher FGR recycle rate would be preferred.

6.5 THE EFFECT OF OPERATING MLSS CONCENTRATION ON PROCESS PERFORMANCE

The investigation of the effects of steady-state operating MLSS concentration was carried out during Phase 5, when the average daily volume of mixed liquor wasted from the aeration basin was 115 L on the A Side, compared to 195 L on the B Side. The resulting aeration basin mean MLSS concentration was 3089 mg/L on the A Side, compared to 1965 on the B Side.

The t test comparisons of mass balances conducted on the individual process reactors during Phase 5 showed that, because of the low ammonia loading to the FGRs, the lower operating MLSS concentration on the B Side did not significantly affect the degree of ammonia removal or nitrate production over the entire process. No significant differences between the A and B Sides were detected in either mean ammonia removal or mean nitrate production in FGR Cells 1 and 2, in the aeration basin, or across the entire process. Similarly, nitrate reduction in the anoxic reactor did not differ significantly between the two sides (see Tables 10 and 11). However, the batch test results showed that the suspended growth nitrification and

denitrification rates were lower on the B Side than the A Side during this period (see Figure 8).

The batch test data for Phase 5 are compared in Table 22. As shown, the suspended growth rates of NO_x reduction in the anoxic phase were higher on the A Side (5.0-6.2 mg N/L) than on the B Side (3.7-3.8 mg N/L). However, the specific rates of anoxic NO_x reduction were equivalent on both sides (1.6-1.7 mg N/hr/g MLSS). Therefore, it is apparent that the higher NO_x reduction rate on the A Side was simply due to the greater concentration of MLSS, and not to any difference in the bacterial population (since there was little time for acclimation, it is not surprising that changes in the bacterial population did not occur). Similarly, the rates of NO_x production in the aerobic phase were higher on the A Side (3.1 mg N/L/hr) than on the B Side (1.7-1.8 mg N/L/hr), but the specific NO_x production rates were equivalent on both sides (0.8-1.0 mg N/hr/g MLSS).

Table 22 Batch Test Comparison of the Effects of Operating MLSS Concentration on Denitrification, Nitrification, and Phosphorus Removal Rates - Phase 5

Parameter	Data Source	Uptake or Release Rate (mg/L/hr)		Specific Uptake or Release rate (mg/hr/g MLSS)	
		A Side	B Side	A Side	B Side
Anaerobic VFA Uptake Rate	Batch 1	60.5	17.1	27.7	9.9
	Batch 2	74.8	30.0	31.0	18.9
Anaerobic PO_4^{3-} Release Rate	Batch 1	40.0	15.2	18.3	8.8
	Batch 2	58.9	13.2	24.4	8.4
Anoxic NO_x Reduction Rate	Batch 1	5.0	3.8	1.6	1.6
	Batch 2	6.2	3.7	1.7	1.7
Aerobic NO_x Production Rate	Batch 1	3.1	1.8	1.0	0.8
	Batch 2	3.1	1.7	0.8	0.8
Aerobic PO_4^{3-} Uptake Rate (overall average)	Batch 1	5.0	2.3	1.6	0.9
	Batch 2	5.6	1.9	1.5	0.9

As discussed earlier, nitrification in the pilot plant was mainly accomplished in the FGRs, and very little ammonia reached the aeration basin (Chapter 5). However, in a plant heavily loaded with ammonia, where the flow leaving the FGR(s) might have significant concentrations of ammonia, the higher suspended-growth nitrification rate observed on the A Side (average operating MLSS=3090 mg/L) compared to the B Side (average operating MLSS=1970 mg/L) could have a significant effect on effluent quality.

The rates of anaerobic VFA uptake and aerobic phosphorus uptake for the batch tests are also compared in Table 22. As shown, the rates of VFA uptake were higher on the A Side (60.5-74.8 mg HAc/L/hr) than on the B Side (17.1-30.0 mg HAc/L/hr). In this case, the specific VFA uptake rates were also higher on the A Side (27.7-31.0 mg HAc/hr/g MLSS) than on the B Side (9.9-18.9 mg HAc/hr/g MLSS). Similarly, in the aerobic phase on the A Side, both the rates of orthophosphate uptake and specific orthophosphate uptake were higher on the A Side (5.0-5.6 mg P/L/hr and 1.5-1.6 mg P/hr/g MLSS, respectively), compared to the B Side (1.9-2.3 mg P/L/hr and 0.9 mg P/hr/g MLSS, respectively). Note that the percent phosphorus by weight in the aeration basin MLSS was significantly higher on the A Side (4.6%) than on the B Side (3.4%), during Phase 5 (Table 13).

The batch test results discussed above are supported by the results of pilot plant monitoring. As discussed earlier, the difference in operating MLSS concentration between the two sides had a significant effect on process phosphorus removal. Mean release of orthophosphate and mean specific release of orthophosphate in the A Side anaerobic reactor were significantly higher than on the B Side, and the corresponding mean mass of anaerobic VFA uptake was also significantly higher on the A Side (Tables 12 and 14). Under aerobic conditions in FGR Cell 1, bacterial uptake of orthophosphate was significantly higher on the A Side (mean=14.8 mg/L) than the B

Side (mean=4.3 mg P/L - see Table 14). Similarly, mean orthophosphate uptake in the A Side FGR Cell 2 (8.8 mg P/L) and the A Side aeration basin (7.1 mg P/L) were significantly higher than on the B Side (4.7 mg P/L and 5.3 mg P/L for FGR Cell 2 and the aeration basin, respectively). Overall phosphate uptake across the anoxic reactor, FGR Cells 1 and 2, and the aeration reactor was 38.4 mg P/L for the A Side, compared to only 20.9 mg P/L on the B Side (Table 14). The above results are consistent with the results of full-scale studies at Salmon Arm, where plant effluent phosphorus concentration increased as the mean aeration basin operating MLSS concentration was reduced from 4090 mg/L to 3250 mg/L, and then to 2360 mg/L; however, note that the Salmon Arm study did not involve parallel comparisons between a control and experimental process, and the comparisons among different operating MLSS concentrations may have been influenced by uncontrolled seasonal environmental variations (Gibb, 1990).

The results discussed above for nitrification during Phase 5 suggested that the lower nitrification rate associated with the lower operating MLSS concentration was simply due to a lower mass of solids in the aerobic phase, and not to any obvious difference in the percentage of nitrifiers in the MLSS, since the specific nitrification rate was similar for both sides. On the other hand, the lower operating MLSS concentration on the B Side resulted in a significantly lower phosphorus content in the MLSS compared to the A Side, as well as lower specific rates of anaerobic VFA uptake-phosphorus release, and aerobic phosphorus uptake on the B Side. One possibility is that the lower sludge age on the B Side resulted in a reduction in the percentage of phosphorus accumulating bacteria in the suspended growth, due to washout. According to Wentzel et al. (1991a and 1991b), phosphorus accumulating bacteria have an abnormally low endogenous decay rate; longer sludge ages in activated sludge BNR systems therefore might lead to an increase in the proportion of phosphorus accumulating bacteria in the active mass, with a resultant increase in the

phosphorus content of the MLSS.

A second possibility is that the lower MLSS concentration on the B Side resulted in a lower steady-state phosphorus accumulating capacity of individual bacteria on that side. As discussed earlier in Section 6.2.3, VFA uptake on the B Side during Phase 5 was apparently limited by the anaerobic actual HRT, as evidenced by the high steady-state VFA concentration (4.51 mg HAc/L) in the anaerobic reactor. At the outset of Phase 5, the lower VFA uptake on the B Side would simply be due to insufficient HRT for the lower mass of MLSS to take up all available VFA. However, over time, the lower degree of anaerobic carbon storage due to HRT limitations would result in a decline in the phosphorus content of the biomass, which in turn would cause a decrease in the anaerobic VFA uptake rate, resulting in a further drop in the mass of anaerobic carbon storage, and so forth. If the opportunity for carbon storage on the B Side during Phase 5 had been increased by increasing the anaerobic actual HRT (say to 45 minutes), phosphorus removal on the B Side might have been equivalent to that on the A Side. In this scenario, the deterioration in phosphorus removal on the B Side during Phase 5 would be associated with sludge age only to the extent that the lower mass of MLSS caused VFA uptake in the process to become limited by the anaerobic HRT, and not to any inherent change in the bacterial population. This is the more likely case, since biological phosphorus removal has been shown to be effective at sludge ages ranging from three to sixty days (Randall et al., 1992). As noted earlier, a longer period of acclimation might have resulted in an increase in the percentage of phosphorus accumulating bacteria in the process MLSS on the B Side, with a consequent improvement in phosphorus removal.

6.6 BATCH TEST SIMULATIONS OF MANIPULATING PROCESS RECYCLE RATES

6.6.1 Denitrified Recycle Rate

The results of batch test simulations to investigate the single-cycle effects of manipulating the denitrified recycle rate to gain additional actual HRT in the anaerobic reactor and to attenuate the dilution of anaerobic MLSS concentration in response to a short-term increase of 65% in process influent flow rate were presented in Section 5.2.6. The full-scale studies at Salmon Arm showed that diurnal fluctuations in process influent flow rate could have a significant impact on process phosphorus removal (Gibb, 1994).

The batch test results showed that reducing the denitrified recycle rate to buffer the effects of hydraulic shocks caused by increases in process influent flow rate would not necessarily be an effective maneuver. The increase in anaerobic HRT gained by simulating a reduction in the denitrified recycle appeared to be more than offset by a decrease in anaerobic MLSS concentration and an increase in anaerobic initial VFA concentration, which resulted in incomplete bacterial uptake-storage of available VFA in the anaerobic phase. Allowing significant concentrations of VFA to reach the anoxic phase resulted in a decrease in net phosphate uptake during denitrification in the anoxic reactor, increasing the phosphorus loading to the aerobic phase. As noted earlier, the batch tests were based on a single short-term event; repeated daily peak loads might eventually result in a change in process behavior, due to acclimation of the biomass to the repeated cycle of daily peak loading.

On the other hand, simulating an increase in the denitrified recycle reduced the dilution of the anaerobic MLSS, reduced the anaerobic initial VFA concentration, and resulted in faster removal of available VFA during the anaerobic phase. According to theory, increased anaerobic carbon storage should result in a greater degree of

phosphate uptake in the subsequent aerobic phase. However, this was not the case in the batch tests. In spite of the increase in carbon storage realized by increasing the denitrified recycle rate, the bulk solution orthophosphate concentration in Reactors #4 and #5 was greater than 1 mg P/L, compared to only 0.55 mg P/L in the control reactor. It may be that the effects of increased (or decreased) anaerobic carbon storage on biological phosphorus removal only become apparent over a large number of cycles through the plant. Some researchers have found that the effects of changes in organic loading to BNR plants take up to several days to become apparent (e.g., Meganck et al., 1985)

As described in Chapters 4 and 5, the above batch test was not designed to reflect the gradual changes in reactor MLSS concentrations which would result from fluctuations in process influent flow rate and denitrified recycle flow rate. However, the batch test results did provide a general indication of the short-term effects of manipulating the denitrified recycle rate in response to an increase in process influent flow rate. The results indicated that the most effective strategy in a full-scale plant would be to leave the recycle rate at its steady-state value, provided that the anaerobic HRT was adequate to allow bacterial uptake of all available VFA (assuming that the full-scale plant were configured in a similar way to the pilot plant). In a plant where the reduction in anaerobic HRT caused by the peak daily load resulted in incomplete bacterial uptake of all available VFA in the anaerobic phase, it might be beneficial to increase the denitrified recycle rate during the peak flow period, to increase the degree of carbon storage. Whether this would result in improved phosphorus removal over the long term is unknown; however, current theory suggests that this should be the case.

6.6.2 Settled Biosolids Recycle Rate

The results of batch test simulations to investigate the single-cycle effects of

manipulating the return settled biosolids recycle rate to gain additional actual HRT in the anoxic and aerobic suspended growth reactors and to attenuate the dilution of anoxic and anaerobic MLSS concentration in response to an increase of 65% in process influent flow rate were presented in Section 5.2.7. The poorest performance in terms of phosphorus removal was observed in the control reactor, where the recycle was maintained at its steady-state value of Q, and final orthophosphate concentration was 4.6 mg P/L. Simulating a decrease in the recycle to 0.5Q and 0.25Q to gain additional aerobic HRT improved phosphorus removal to final orthophosphate concentrations of 4.0 mg P/L and 3.4 mg P/L, respectively. On the other hand, increasing the biosolids recycle to 1.5Q and 2Q to reduce dilution of MLSS concentration also improved phosphorus removal, to final concentrations of 4.1 and 3.3 mg P/L, respectively.

As described in Chapters 4 and 5, the above batch test was not designed to reflect the gradual changes in reactor MLSS concentrations which would result from fluctuations in process influent flow rate and the returns settled biosolids flow rate. However, the batch test results did provide a general indication of the short-term effects of manipulating the return biosolids flow rate in response to an increase in process influent flow rate. The results indicated that the settled biosolids recycle rate has potential as a process control parameter to help deal with the daily peak hydraulic loads at full-scale plants, both through decreasing the recycle to gain additional HRT and through increasing the recycle to attenuate dilution of the process MLSS. However, it should be noted that decreasing the biosolids recycle rate in a UCT-type process might induce secondary phosphorus release in the anoxic reactor, due to the decrease in anoxic NO_x loading.

6.7 SIGNIFICANCE OF RESULTS

As described in the preceding sections and in the following conclusions, this study provided significant insights into the importance of suspended growth HRTs, internal recycle flow rates, and operating MLSS concentration in the FGR-SGR process. Useful design and operating information derived from the study includes the potentially negative consequences of inadequate anaerobic HRT and of excessive anoxic HRT, the importance of regular hydraulic scouring of the FGR media to prevent solids accumulation, the potential to dampen the effects of diurnal flow increases by manipulating internal recycle streams, and the importance of the MLSS concentration in the suspended growth train.

The potential benefit of including a fixed growth component in BNR processes was also demonstrated. The size (HRT) of the aeration basin in activated sludge BNR plants is typically limited by the suspended growth nitrification rate, which is relatively slow (e.g., the batch tests carried out during Phase 3 of this study showed that complete nitrification of ammonia by the suspended growth took approximately twice as long as did phosphorus uptake - see Figure 6). By accomplishing a significant degree of fixed growth nitrification in the FGR (up to 68% - see Table 20), the aeration basin could be reduced to that which is required for phosphorus uptake only. In this case, the required aeration basin size (HRT) could be reduced by approximately 50%, without causing an increase in effluent ammonia or phosphorus concentrations, provided that at least 50% of the ammonia in the process influent were oxidized by the fixed growth upstream of the aeration basin. In cases where an existing trickling filter (FGR) plant is to be upgraded for BNR, incorporation of the FGR into the process train could then result in significant savings in the capital and operating costs of the aeration basin, compared to de-commissioning the FGR and constructing an activated sludge plant with a much larger aeration basin.

7. CONCLUSIONS

As described in Section 2, the objectives of this study were to investigate the optimum hydraulic retention times (HRTs) in the suspended growth components of the FGR-SGR process, to investigate the optimum internal recycle flow rates within the process, and to investigate the optimum system operating mixed liquor suspended solids concentration. The following conclusions are based on the results discussed in Chapters 5-7 of this report.

- The pilot-scale study confirmed the application of the FGR-SGR process to biological nutrient removal. The process produced an effluent with non-detectable ammonia concentrations, and mean orthophosphate concentrations of 0.10-0.27 mg P/L, under optimum conditions.

- Process effluent phosphorus concentration depends on the ratio of VFA taken up in the anaerobic reactor to process phosphorus loading (VFA uptake:influent P); where the ratio is greater than 6 mg VFA (as HAc) taken up per mg influent phosphorus (as P), effluent phosphorus concentrations are generally less than 0.5 mg P/L.

- The FGR-SGR process can operate effectively for biological nutrient removal with a total nominal hydraulic retention time (*i.e.*, total process volume divided by process influent flow rate) of eight hours or less. According to the results of bench-scale batch simulations, the total nominal hydraulic retention time could be reduced to five hours, without compromising effluent quality. Optimum actual hydraulic retention times (*i.e.* process reactor volume divided by process influent flow rate plus any recycle flow through rates) were 8-25 minutes for the anaerobic reactor, and 35 minutes for the anoxic reactor (based

on pilot plant performance and bench-scale batch tests). The optimum actual hydraulic retention time for the aeration reactor was approximately 80 minutes (based on bench-scale batch tests only).

- In cases where the mass of VFA taken up in the anaerobic reactor is limited by the HRT, the optimum anaerobic HRT depends on the process phosphorus loading and the steady-state MLSS concentration; the anaerobic HRT required for adequate VFA uptake is lower at a higher MLSS concentration (based on the results of one month of operation with no preceding acclimation period).

- An increase in the steady-state VFA concentration in the process influent results in a higher VFA uptake rate in the anaerobic reactor, reducing the anaerobic HRT required to achieve an high enough VFA uptake:influent P ratio to result low effluent phosphorus concentrations. However, the higher influent VFA concentration can result in high VFA concentrations reaching the anoxic reactor (see next point).

- The HRT in the anaerobic reactor should be sufficient to ensure bacterial uptake of all available volatile fatty acids (VFA) in the process influent. Allowing high concentrations of VFA to reach the anoxic reactor reduces the net anoxic phosphorus uptake by inducing phosphorus release.

- Bacterial uptake and storage of phosphorus in the FGR-SGR process using nitrates as an electron acceptor is feasible in systems where significant nitrification occurs. The use of nitrates rather than dissolved oxygen for phosphorus removal can reduce the required aeration volume, with a resulting decrease in the energy required for aeration. However, the total volume required for phosphorus uptake (anoxic and aerobic) would not be reduced (others have reported that unit PHA consumption per unit of phosphorus accumulated is lower using nitrate as an electron acceptor, compared to using

oxygen).

- The HRT of the anoxic reactor should be limited to that which is required for complete denitrification of nitrates carried into the reactor in the internal recycle streams. Otherwise, post-denitrification release of phosphorus will result in an increase in the phosphorus loading to the aerobic phase, which in turn will require an increase in the size of the aeration reactor, to allow extra time for bacterial uptake of the additional phosphorus released in the anoxic reactor (provided that bacteria have enough PHA reserves to allow uptake of the phosphorus released in the anoxic zone).

- Short-term daily increases in the FGR recycle rate to hydraulically scour the media and prevent solids accumulation greatly improves phosphorus removal in the FGR-SGR process, reduces fluctuations in operating mixed liquor suspended solids concentration, and is an effective measure for controlling filter flies (*psychoda*).

- Increasing the denitrified recycle rate in response to a short -term increase in process influent flow rate can decrease the time required for completion of bacterial VFA uptake in the anaerobic reactor, by attenuating the dilution of anaerobic MLSS (based on bench-scale batch tests only).

- Increasing the settled biosolids recycle rate in response to a short -term increase in process influent flow rate can increase phosphorus uptake, by attenuating the dilution of anoxic and aerobic MLSS. On the other hand, decreasing the settled biosolids recycle rate in the same situation can also improve phosphorus uptake, by increasing the actual aerobic HRT (based on bench-scale batch tests only).

- Biological phosphorus removal, denitrification, and suspended-growth nitrification rates in the FGR-SGR process are greater at an operating aeration

basin MLSS concentration of 3000 mg/L than at 2000 mg/L (based on the results of one month of operation with no preceding acclimation period).

- Complete nitrification of ammonia by suspended bacteria in the aeration basin generally required significantly longer than complete uptake and storage of phosphorus. Therefore, the addition of the fixed growth component is beneficial, since partial or complete nitrification of ammonia by fixed growth in the FGRs upstream of the aeration basin allows the aeration basin to be sized for phosphorus removal only.

8. RECOMMENDATIONS FOR FURTHER RESEARCH

- Batch test investigations showed that single-cycle manipulations of the denitrified recycle flow rate and the return settled biosolids flow rate have the potential to dampen the effects of the diurnal flow fluctuations typically experienced at full-scale treatment plants. Further work is required to investigate the long-term effects of manipulating the recycle flow rates over a large number of cycles.

- The use of nitrates rather than dissolved oxygen for bacterial uptake and storage of phosphorus in biological nutrient removal systems should be further explored, to determine the relative amounts of PHA consumption to phosphorus uptake.

- The nitrification rate of the FGRs (trickling filters) in the FGR-SGR process should be investigated at ammonia concentrations in excess of the so-called "first order threshold" (i.e., greater than 3-5 mg N/L).

- Other configurations and densities of FGR media should be tested in the FGR-SGR process, to investigate the propensity for plugging of the media by suspended solids, and to determine loading and performance criteria.

- The frequency and duration of hydraulic pulse loading of the FGRs to prevent solids accumulation on the media should be further investigated, to determine the optimum irrigation schedule for different types of media.

- The long-term effects of operating mixed liquor suspended solids concentration on biological phosphorus removal in the FGR-SGR process should be further investigated, to determine whether acclimation of the biomass can offset the effects of a lower MLSS concentration.

- The effects on phosphorus removal of allowing significant concentrations of volatile fatty acids to reach the anoxic reactor should be further investigated, to determine whether the ratio of phosphorus released to PHA stored in the anoxic zone is similar to that in the anaerobic zone.

- The conclusions drawn from batch tests regarding optimization of the hydraulic retention time in the aeration basin should be confirmed through further pilot-scale investigations.

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APPENDIX 1 - PILOT PLANT REACTOR MASS BALANCE CALCULATIONS

A schematic of the pilot plant FGR-SGR process train is shown in Figure 1 in Chapter 4 of this thesis. The mass balance calculations below are referenced to Figure 1, and are normalized to reflect nominal hydraulic retention time (HRT), rather than actual HRT.

Q_O = Primary Clarifier Overflow Flow Rate (L/min)

Q_d = Denitrified Return Flow Rate from Anoxic Reactor to Anaerobic Reactor (L/min)

Q_r = Settled Biosolids Return Flow Rate from Secondary Clarifier Underflow to Anoxic   Reactor

C_O = Concentration in Primary Clarifier Overflow (mg/L)

C_{anaer} = Concentration of Substrate X in Anaerobic Reactor Mixed Liquor (mg/L)

C_{anox} = Concentration of Substrate X in Anoxic Reactor Mixed Liquor (mg/L)

C_{fgr1} = Concentration of Substrate X in FGR 1 Catchbasin Sump Mixed Liquor (mg/L)

C_{fgr2} = Concentration of Substrate X in FGR 2 Catchbasin Sump Mixed Liquor (mg/L)

C_{aer} = Concentration of Substrate X in Aeration Reactor Mixed Liquor (mg/L)

C_r = Concentration of Substrate X in Return Flow from Secondary Clarifier Underflow   to Anoxic Reactor (mg/L)

C_{eff} = Concentration of Substrate X in Secondary Clarifier Overflow (mg/L)

Appendix 1

Normalized Anaerobic Reactor Mass Balance

$$\text{Anaerobic Substrate Removal (mg/L)} = \frac{(C_o Q_o + C_{anox} Q_d) - C_{anaer}(Q_o + Q_d)}{Q_o}$$

Normalized Anoxic reactor Mass Balance

$$\text{Anoxic Substrate Removal (mg/L)} = \frac{(C_{anaer}(Q_o + Q_d) + (C_r Q_r)) - C_{anox}(Q_o + Q_d + Q_d)}{Q_o}$$

Normalized FGR Cell 1 Mass Balance

$$\text{FGR Cell 1 Substrate Removal (mg/L)} = \frac{(C_{anox} - C_{fgr1})(Q_o + Q_d)}{Q_o}$$

Normalized FGR Cell 2 Mass Balance

$$\text{FGR Cell 2 Substrate Removal (mg/L)} = \frac{(C_{fgr1} - C_{fgr2})(Q_o + Q_d)}{Q_o}$$

Normalized Aeration Basin Mass Balance

$$\text{Aeration Basin Substrate Removal (mg/L)} = \frac{(C_{fgr2} - C_{aer})(Q_o + Q_d)}{Q_o}$$

Normalized Secondary Clarifier Mass Balance

$$\text{Secondary Clarifier Substrate Removal (mg/L)} = \frac{(C_{effl} Q_o + C_r Q_r) - C_{aer}(Q_o + Q_r)}{Q_o}$$

APPENDIX 2 - METHOD OF CALCULATING VOLUME OF SAMPLE ADDED FOR BATCH TEST INVESTIGATIONS OF DENITRIFIED RECYCLE RATE AND SETTLED BIOSOLIDS RECYCLE RATE

The anaerobic, anoxic, and aerobic phases of the batch tests were conducted to match the reactor sequence in the pilot plant as closely as possible. The following calculation is for Reactor #1 in the investigation of the denitrified recycle rate (Table 4). The other values for volume of sample added shown in Tables 4 and 5 followed the same method.

For Reactor #1 in the investigation of denitrified recycle rate, the simulated flow rates were as follows:

process influent flow rate = 1.65Q
denitrified recycle flow rate = 2Q
return settled biosolids flow rate = Q

1. Anaerobic Phase

A sample of process influent was taken from the primary clarifier overflow pipe, and a sample of denitrified mixed liquor was taken from the anoxic reactor near the overflow weir.

total volume of batch reactor = 2,800 mL
total simulated flow into anaerobic reactor (see Figure 1)= 1.65Q + 2Q = 3.65Q

$$\text{volume of process influent added at time zero} = \frac{(1.65Q)(2,800 \text{ mL})}{3.65Q} = 1,265 \text{ mL}$$

$$\text{volume of anoxic mixed liquor added at time zero} = \frac{(2Q)(2,800 \text{ mL})}{3.65Q} = 1,535 \text{ mL}$$

Appendix 2

2. Anoxic and Aerobic Phase

Since the mixed liquor from the pilot plant anaerobic reactor flowed into the anoxic reactor to meet the return settled biosolids stream, the sample taken from the secondary clarifier underflow return line was added to the mixed liquor remaining in the batch reactor at the end of the anaerobic phase. Therefore, enough mixed liquor had to be removed from the batch reactor during the anaerobic phase to make room for the sample of return settled biosolids.

total simulated flow into anoxic reactor (see Figure 1) = 1.65Q + 2Q + Q = 4.65Q

$$\text{required volume of mixed liquor left in batch reactor at end of anaerobic phase (i.e., at beginning of anoxic phase)} = \frac{(1.65Q+2Q)(2{,}800 \text{ mL})}{4.65Q} = 2{,}200 \text{ mL}$$

$$\text{volume of return settled biosolids added at end of anaerobic phase (i.e., at beginning of anoxic phase)} = \frac{Q(2{,}800 \text{ mL})}{4.65Q} = 600 \text{ mL}$$