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Optimization of Biological Phosphorus and Ammonia Removal in a Combined Fixed and Suspended Growth Wastewater Treatment System

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FINAL REPORT FOR:

OPTIMIZATION OF BIOLOGICAL PHOSPHORUS AND AMMONIA REMOVAL IN A COMBINED FIXED AND SUSPENDED GROWTH WASTEWATER TREATMENT SYSTEM

DECEMBER, 1994

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SUMMARY

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.

In the FGR-SGR process, nitrification is mainly accomplished by the fixed growth biomass attached to the FGR (trickling filter) media. The long aeration basin hydraulic retention times required for nitrification in activated sludge systems are therefore unnecessary in the FGR-SGR process, and the aeration basin size may be optimized for bacterial phosphorus uptake only. Other advantages of the FGR-SGR process include the option of retro-fitting existing FGR facilities for biological phosphorus and nitrogen removal, low energy cascade aeration of the process mixed liquor, process stability added by the fixed growth biomass, reduced competition among the specific microorganisms responsible for uptake-storage of phosphorus and nitrification due to separation of the respective unit processes into distinct microbial communities, and reduction of the land area required for full-scale biological phosphorus removal facilities.

The pilot-scale study was designed to extend and refine the design and operational criteria for biological phosphorus and nitrogen removal for the next generation of full-scale FGR-SGR plants. The objectives of the pilot-scale study were to investigate the optimum hydraulic retention times in the activated sludge reactors, the optimum internal recycle flow rates within the process, and the optimum solids retention time for the suspended growth treatment train.

Two pilot-scale processes were operated in parallel, to compare plant performance in response to controlled changes in design and operating parameters. Bench-scale batch tests designed to simulate the suspended growth treatment train were periodically conducted using grab samples of the process mixed liquor, to gain additional insight into biochemical reaction rates.

The optimization scheme resulted in a reduction in system total nominal hydraulic retention time (process influent flow rate divided by total process

volume) from approximately 12 hours to less than 8 hours. Results indicated that the process total nominal retention time could be reduced to as little as 5 hours, without compromising effluent quality.

During the course of the study, the anaerobic reactor was reduced from an actual hydraulic retention time (process volume divided by process influent flow rate plus any recycle flow rates) of 45 minutes to 8 minutes. The results showed that the process could achieve biological phosphorus removal to a mean effluent orthophosphate concentration of 0.27 mg P/L with the actual anaerobic hydraulic retention time (HRT) of 8 minutes, provided that the concentration of volatile fatty acids (VFA) in the process influent was in the range 30-50 mg/L as acetic acid.

The actual HRT of the anoxic reactor was reduced from 65 minutes at the beginning of the study to an optimum value of 35 minutes. Study of the anoxic reactor showed that denitrification of nitrate was accompanied by bacterial uptake of phosphorus. After all the nitrate was removed from solution, secondary phosphorus release was observed. The results demonstrated that if the anoxic reactor is oversized (or, alternatively, under loaded with nitrate), 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.

The results of bench-scale batch tests indicated that the actual HRT in the aeration reactor could be reduced from the initial value of 160 minutes to approximately 80 minutes, without compromising biological phosphorus removal. The time required for complete nitrification of ammonia by suspended bacteria in the aeration reactor was generally longer than the time required for complete bacterial removal of phosphorus. However, in the FGR-SGR process, most of the nitrification was accomplished upstream of the aeration reactor, by fixed growth bacteria attached to the FGR media, indicating that the aeration reactor HRT could be optimized for phosphorus removal only. Unfortunately, time constraints and the need to complete the other study tasks did not allow operation of the pilot plant at different aeration basin sizes, and conclusions regarding the aerated HRT were based on the results of bench-scale batch tests only.

In the FGR-SGR system, the process mixed liquor was irrigated over the FGR media. In the early phases of the study, the FGR irrigation rate was always maintained at a steady-state value. After approximately one year of operation, it was found that excessive solids were building up on the media. At the same time, a significant increase in effluent phosphorus concentration was observed. A system of daily hydraulic pulse loading of the FGRs was subsequently instituted, to scour excess solids from the media. Following the beginning of the pulse loading regime, biological phosphorus removal to low effluent concentrations was immediately restored. Hydraulic pulse loading of the FGRs also reduced fluctuations in MLSS concentration in the suspended

growth treatment train, and virtually eliminated filter fly (*psychoda*) infestations.

For most of the study, the FGR recycle (irrigation) rate was maintained at 24 L / min., except for the daily pulse increase to scour excess solids from the media. The study results showed that reducing the FGR recycle rate to 14 L / min did not result in a significant deterioration in process phosphorus removal or nitrification. The lower recycle rate was therefore determined to be the optimum, since it would result in lower pumping costs at a full-scale plant. A recycle rate lower than 14 L / min was not investigated, since it would have resulted in incomplete wetting of the media.

Time constraints did not allow operation of the pilot plant to investigate the effects of manipulating the denitrified recycle flow rate or the return biosolids flow rate. The purpose of manipulating the recycle flow rates would be to dampen the effects of the peak daily hydraulic load typically experienced at a full-scale plant. However, batch test simulations were developed to investigate the effects of manipulating the recycle flows over a single cycle through the plant.

According to the results of the bench-scale simulations, manipulation of the denitrified recycle rate (*i.e.*, the mixed liquor return flow from the anoxic reactor to the anaerobic reactor) in response to an increase in process influent flow rate had a detrimental effect on phosphorus removal over a single cycle. Simulation of a decrease in the denitrified recycle indicated that the resulting gain in actual anaerobic HRT was more than offset by the associated dilution of mixed liquor suspended solids in the anaerobic reactor; the end result was incomplete bacterial uptake of available VFA in the anaerobic phase. However, simulation of an increase in the denitrified recycle resulted in faster bacterial removal of VFA in the anaerobic phase. Apparently, the decrease in anaerobic HRT resulting from the higher recycle rate was more than overcome by the associated increase in anaerobic MLSS concentration. It follows that increasing the denitrified recycle rate to maintain a higher operating MLSS concentration in the anaerobic reactor during the peak daily hydraulic load might improve biological phosphorus removal over a large number of cycles, by increasing the degree of anaerobic carbon storage.

According to the results of bench-scale simulations, manipulation of the settled biosolids recycle rate (*i.e.*, the return flow of settled biosolids from the secondary clarifier to the anoxic reactor) to dampen the effects of an increase in process influent flow rate has the potential to improve phosphorus removal in the process. The simulations indicated that the additional aerobic HRT gained from a decrease in biosolids recycle can result in increased phosphorus removal, compared to maintaining the recycle at its steady-state value. On the other hand, the simulations indicated that the higher operating MLSS concentration in the aerobic reactor resulting from an increase in biosolids recycle can also improve phosphorus removal during the peak daily flow, compared to maintaining the steady-state recycle rate. According to the batch

test results, increasing the recycle by 50% would have approximately the same beneficial effect as decreasing the recycle by 50%. However, decreasing the biosolids recycle would be the better course at a full-scale plant, since increasing the recycle might overload the final clarifier.

The investigation of optimum solids retention time (SRT) in the suspended growth treatment train showed that biological phosphorus removal was significantly better at a longer SRT. At a secondary solids wasting rate which resulted in a mean operating MLSS concentration in the aeration basin of 3090 mg/L, mean phosphorus removal from the process influent was 3.6 mg P/L, compared to a mean phosphorus removal of only 1.8 mg P/L at a wasting rate which resulted in a mean aeration basin MLSS concentration of 1970 mg/L. Suspended growth nitrification rates and denitrification rates were also significantly higher at the higher operating MLSS concentration.

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 the design and operating configuration. Under optimized conditions, mean effluent total phosphorus and orthophosphate concentrations were in the range 0.25-0.35 mg P/L and 0.10-0.27 mg P/L, respectively. Samples of mixed liquor suspended solids taken from the aeration basin typically contained approximately 4% phosphorus by dry weight, indicating a significant degree of excess bacterial storage of phosphorus.

The study results have confirmed the capacity of the FGR-SGR process for effective biological phosphorus removal and nitrification-denitrification. The results show that trickling filter plants can be retro-fitted to include biological phosphorus removal, eliminating or reducing the requirement for chemical additions, and maximizing the use of existing facilities. The study data are currently being applied to the design of full-scale plants.

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1 INTRODUCTION

The discharge of nutrient-rich wastes such as domestic sewage into environmentally-sensitive surface waters (*e.g.*, Shuswap Lake in the Fraser River Basin) can cause serious a 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 in unionized form is toxic to fish.

Chemical removal of phosphorus is currently practiced in many applications, 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 (*e.g.*, the Bardenpho and University of Cape Town processes); existing trickling filters are usually decommissioned and demolished or converted to other uses when activated sludge systems are installed. Use of trickling filters as a key component in a biological nutrient removal process could, therefore, be beneficial.

Current technology for biological conversion of ammonia to nitrate (*i.e.*, nitrification) is based on 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 conventional activated sludge systems generally requires relatively long hydraulic retention times for the process liquid in the aeration basin; increasing capital and operating costs. Nitrification in activated sludge systems also requires relatively long solids retention times (sludge ages), which in some cases can create a poor-settling biomass, which in turn requires a larger secondary clarifier. According to Randall *et.al.* (1992), typical hydraulic retention times in the aeration basin of conventional activated sludge systems designed for nitrification are in the range 3-6 hours, and solids retention times are in the range 5-15 days, depending on the temperature of the process liquid.

In recent years, a system called the Trickling Filter/Solids Contact (TF/SC) process, which incorporates trickling filters in series with a small activated sludge basin for removal of oxygen demand (and in some cases ammonia), has

been developed and successfully applied at more than fifty locations in the U.S.A; the TF/SC process is reported to be lower in capital, operating, and maintenance costs than competing activated sludge processes (Parker *et. al.*, 1990). Another combination trickling filter-activated sludge system, called the Activated Biofilter process, has been installed in at least 43 locations in the U.S.A. for the removal of oxygen demand (Arora and Umphres, 1987). 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.

In activated sludge biological nutrient removal systems (*e.g.*, Bardenpho, University of Cape Town, etc.), 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.

A biological nutrient removal process combining both trickling filter and activated sludge technologies was pioneered by Dayton & Knight Ltd. of West Vancouver, B.C., at a full-scale domestic wastewater treatment facility at Salmon Arm, British Columbia, in 1986. The process is called the FGR-SGR (fixed growth reactor-suspended growth reactor) process. 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 the removal of oxygen demand and ammonia in combined trickling filter-activated sludge systems (Kelly, 1987).

In the FGR-SGR process, nitrification is mainly accomplished by the fixed growth biomass attached to the FGR (trickling filter) media. The long aeration basin hydraulic retention times required for nitrification in activated sludge systems are therefore unnecessary in the FGR-SGR process, and the aeration basin size may be optimized for bacterial phosphorus uptake only. A more detailed description of the process and its various unit operations is provided in Chapter 3 of this report.

The application of the FGR-SGR process for biological phosphorus removal and nitrification has the following advantages:

- The option of retro-fitting existing FGR (trickling filter) facilities for biological phosphorus and nitrogen removal, without de-commissioning of the FGRs - There are presently at least six full-scale trickling filter plants in British Columbia, approximately 2700 in the United States, and many others in overseas markets.
- Lower aeration costs compared to activated sludge-type systems - The FGR-SGR process relies mainly on cascade aeration of the process liquid

over the FGR media. Activated sludge systems often rely on compressed air diffusers, a relatively energy-intensive process.

- Process stability added by the fixed growth biomass - The entire biomass of activated sludge systems may be "washed out" and lost, due to hydraulic shocks. Several weeks or months may then be required to reestablish the system biomass. The fixed growth component of the FGR-SGR system is not as susceptible to hydraulic shocks; removal of biochemical oxygen demand (BOD) and nitrification of ammonia by the fixed growth bacteria in the film attached to the FGR media will continue even if the suspended growth component is washed out.
- Reduced competition among the specific microorganisms responsible for the uptake-storage of phosphorus and nitrification, due to separation of the respective unit processes into distinct microbial communities - This allows optimization of environmental conditions for different bacterial communities in separate reactors.
- Reduction of the land area required for full-scale biological phosphorus removal facilities.
- Elimination of the long hydraulic retention times associated with nitrification in activated sludge systems.

The optimization of biological phosphorus and nitrogen removal processes, including fixed film versus activated sludge systems, was identified as a Biotechnology Research and Development Priority by the Strategic Planning for Applied Research and Knowledge (SPARK) Technical Advisory Committee, an initiative funded through contributions from the Province of British Columbia and the Government of Canada.

Feasibility studies carried out at the full-scale facility at Salmon Arm during 1988-1989 showed that the FGR-SGR process has the capacity for effective biological phosphorus removal (Gibb *et. al.*, 1989). However, the difficulties associated with full-scale research prevented the development of detailed design and operating criteria required for the next generation of design. Following the study at Salmon Arm, it was concluded that pilot-scale work was the best alternative for developing the needed design and operating data.

The project described here involved pilot-scale research and development work on the FGR-SGR process for biological removal of phosphorus and ammonia nitrogen from sewage. The purpose of the study was to optimize process reactor sizes, internal recycle flow rates, and system solids retention time. Data generated from the study are being applied to design of the next generation of full-scale plants.

This project was directly applicable to the goals of Canada's Green Plan. Successful completion of this study promoted the use of an effective, energy-efficient, low-cost nutrient removal system, and thereby helped to enhance the long-term sustainability of environmentally-sensitive surface waters in Canada

(particularly in the Fraser River Basin), by protecting such water uses as recreation, commercial and recreational fisheries, and tourism.

Further, the work increased the knowledge base in British Columbia regarding biological nutrient removal systems. British Columbia is already known internationally as a leader in this field. In addition, publication of the study results will enhance the reputation of Canadian consulting firms, and increase the likelihood of the export of Canadian engineering expertise.

The study results are currently being applied to design of the next upgrade-expansion of the existing trickling filter plant at the Resort Municipality of Whistler, B.C. Other potential sites for application of the study results to retro-fitting and/or upgrading of existing trickling filter plants in B.C. include Salmon Arm, Vernon, Chilliwack, and Abbotsford-Matsqui. The proposed upgrade to secondary treatment by the Greater Vancouver Regional District is based on the trickling filter-solids contact process, which may be upgraded to incorporate the FGR-SGR process with a relatively small investment. There are many other suitable locations for the application of this technology, including the U.S.A., Australia, Sweden, Norway, England, and Africa, and there has been considerable international interest in the project.

This project represented a three-year collaborative effort among B.C. Research, Dayton & Knight Ltd. Consulting Engineers, the University of B.C. Department of Civil Engineering, and the B.C. Institute of Technology. Funding for design and construction of the pilot-scale facility and for the initial phases of the research work was provided by the Science Council of British Columbia, the Fraser Pollution Abatement Office (FPAO) of Environment Canada, and the study collaborators.

The final phases of the research work as included in the technical proposal submitted to the Environmental Innovation Program dated July 15, 1993, entitled "Optimization of Biological Phosphorus and Ammonia Removal in a Combined Fixed and Suspended Growth Wastewater Treatment System" was funded by the Environmental Innovation Program of Environment Canada, under Supply and Services Canada, DSS Contract #KA601-4-0130/01-XSB, the Science Council of B.C., Environment Canada (FPAO), B.C. Research, Dayton & Knight Ltd., and the University of B.C.

The following is the Final Report for DSS Contract #KA601-4-0130/01-XSB. The report contains the results of the work to optimize reactor hydraulic retention times, recycle flow rates, and solids retention time, referred to as Tasks 1, 2, and 3, respectively, in the proposal dated July 15, 1993. Some of the work was completed prior to DSS Contract #KA601-4-0130/01-XSB; however, a summary of all of the results is included in this report for continuity, and to put the more recent results in perspective.

2 OBJECTIVES

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

- Completion of the investigation of optimum hydraulic retention times (*i.e.*, relative reactor sizes) for biological phosphorus removal in both the unaerated and aerated phases of 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 solids retention time (or, alternatively, 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 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 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 substrates to the anaerobic phase which are necessary for enhanced biological phosphorus removal. A diagram of the site layout is shown in Figure 1.

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. A schematic of the pilot-scale FGR-SGR process trains is shown in Figure 2. The process design was developed from the design of the full-scale facility at Salmon Arm, according to the general requirements of biological phosphorus removal as summarized below.

- An anaerobic (no dissolved oxygen present) - aerobic (dissolved oxygen present) sequence is required in the suspended growth bioreactor. According to current biochemical models (*e.g.*, reviewed by Wentzel *et. al.*, 1991a and 1991b), the bacteria responsible for biological phosphorus removal can take up and store soluble carbon in the absence of electron acceptors (*i.e.*, in the anaerobic zone). Uptake and storage of carbon under anaerobic conditions is accompanied by degradation of intracellular phosphate reserves, and the release of orthophosphate ions to the bulk solution. Under subsequent aerobic conditions, the bacteria oxidize the stored carbon, using part of the resulting energy to rebuild intracellular phosphorus 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 solids phases must be separated (usually by gravity settling), and at least some of 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 (*e.g.*, 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 (*e.g.*, Wentzel *et. al.*, 1991a and 1991b).

- 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).

Each pilot-scale 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 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 watertight 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.

Each FGR cell had a cross sectional area of 610 cm x 610 cm, and a height of 5.5 m. The FGR media was 60^0 cross flow plastic, 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 cm intervals and inserted between the media modules, to direct the flow back into the media, and minimize streaming down the cell walls. Sampling ports were located at 1.2 m intervals across 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 through the media cross-section, and process liquid flowing out of the tube was directed into a filter funnel.

Soluble carbon in the settled sewage feed to the anaerobic reactor was supplemented by simple carbon substrates produced by the 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 2). The soluble products of fermentation, mainly acetic acid, propionic acid and butyric acid, were carried to the anaerobic reactor with the clarifier overflow, as the fermented solids resettled along with the influent primary solids. The fermenter design was based on work by Rabinowitz and Oldham (1985).

The anaerobic-anoxic sequence (Figure 2) was modeled on a variation of the activated sludge University of Cape Town (UCT) process. The UCT configuration was selected because it can be designed to ensure a zero discharge of nitrates in the return biological solids stream entering the aerobic reactor (Wentzel *et.al*., 1991a). Settled biological sludge from the final clarifier was returned to the anoxic reactor, where it met the mixed liquor leaving the anaerobic reactor. In the anoxic reactor, carbon-based substrates remaining in the process liquid 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 uptake-phosphorus release (Figure 2).

In diffused air 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. In the FGR-SGR system, the mixed liquor from the anoxic reactor was irrigated over two trickling filters in series (FGR 1 and FGR 2 in Figure 2), and oxygen for phosphorus uptake by suspended organisms was provided by cascade aeration.

Nitrification in trickling filter systems is inhibited by concentrations of soluble five-day BOD greater than 20 mg/L (Parker and Richards, 1986). The FGR-SGR pilot plant was designed for the removal of soluble five-day BOD in the anaerobic and anoxic reactors to concentrations of 20 mg/L or less, to promote nitrification by the fixed growth organisms attached to the FGR media.

Mixed liquor was pumped from the FGR effluent sump to the aeration reactor, for flocculation of sloughed FGR solids and removal of residual phosphorus, ammonia, and BOD by suspended bacteria. Mixed liquor from the aeration reactor flowed to the final clarifier, where the clarified effluent was discharged to sewer, and settled biological solids were returned to the anoxic reactor.

Figure 1- Site Layout

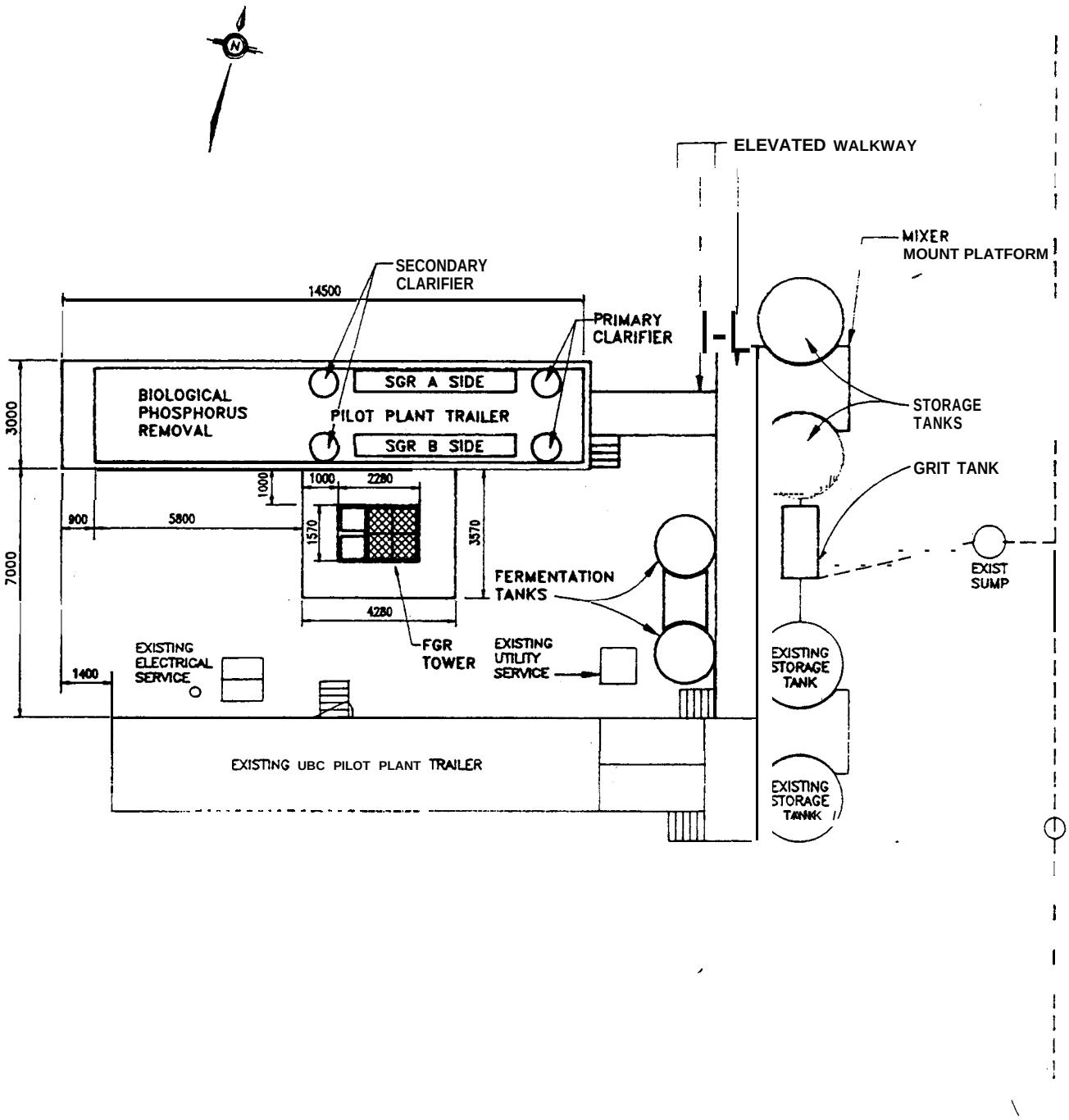
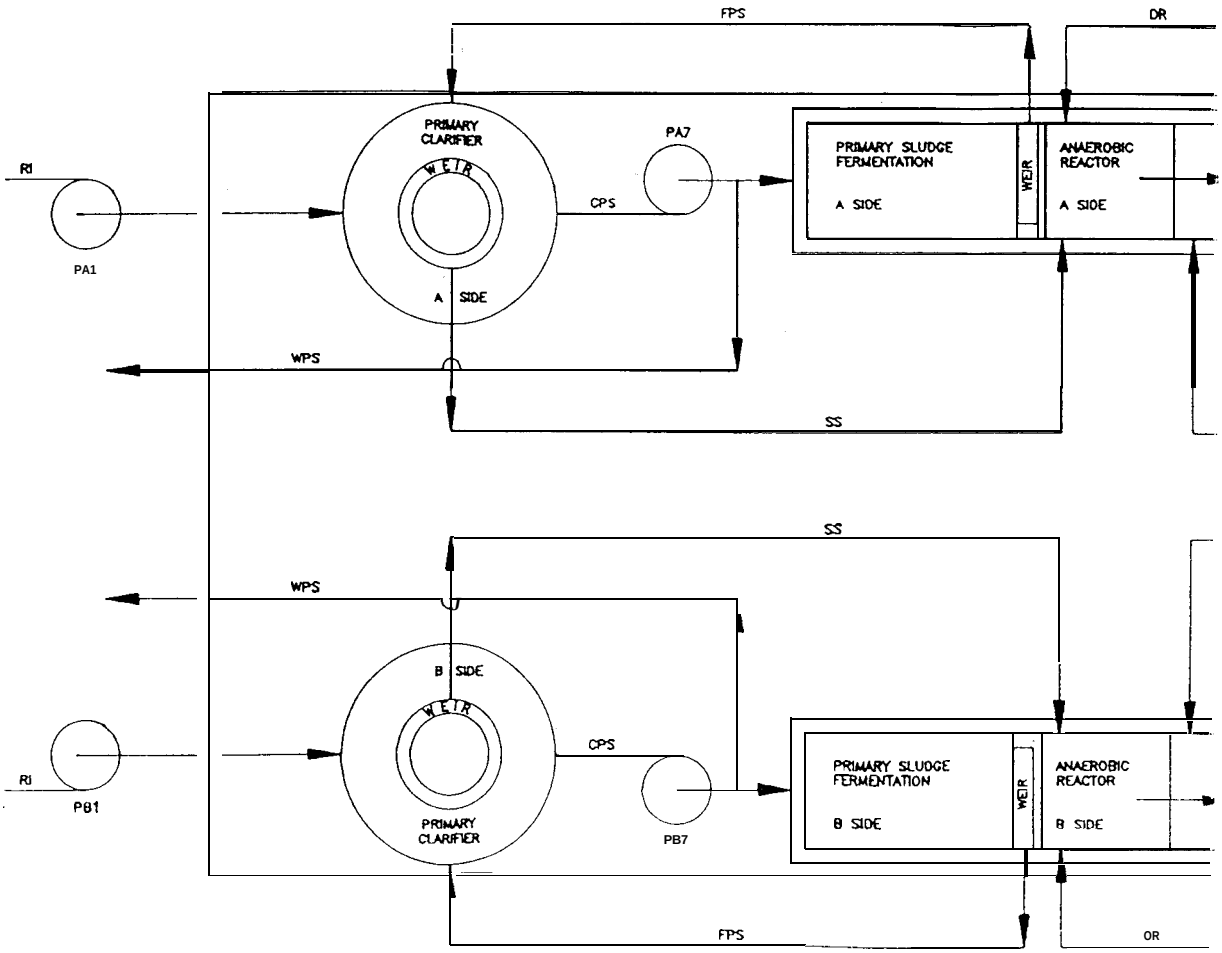
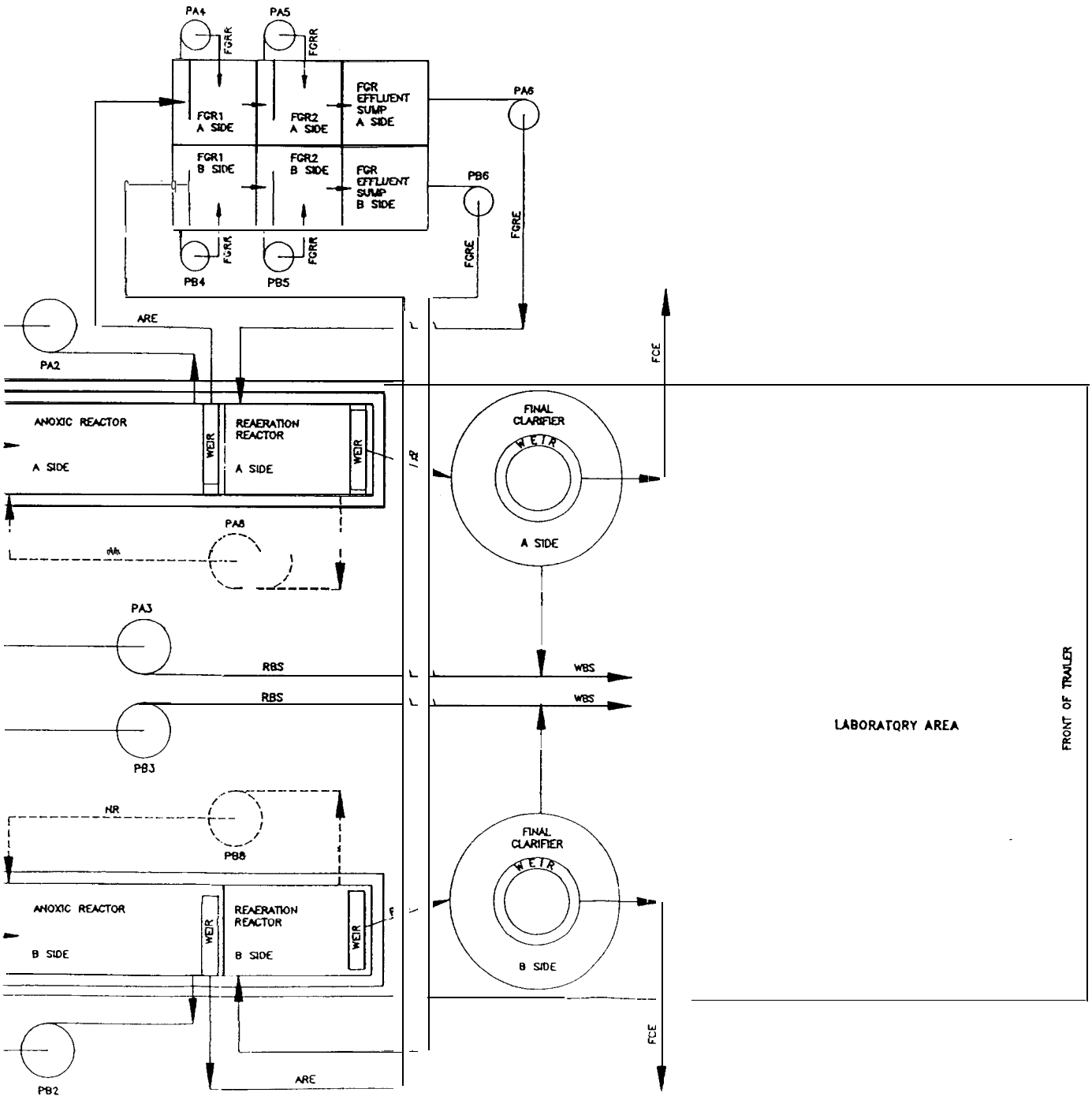


Figure 2- FGR-SGR Pilot Plant Flow Schematic

LEGEND	
RI	RAW INFLUENT
CPS	CRUDE PRIMARY SLUDGE
WPS	WASTE PRIMARY SLUDGE
FPS	FERMENTED PRIMARY SLUDGE
SS	SETTLED SEWAGE
ARE	ANOXIC REACTOR EFFLUENT
RRE	REAERATION REACTOR EFFLUENT
FGRR	FGR RECYCLE
FCE	FINAL CLARIFIER EFFLUENT
WBS	WASTE BIOLOGICAL SLUDGE
RBS	RETURN BIOLOGICAL SLUDGE
OR	DENITRIFIED RECYCLE
NR	NITRIFIED RECYCLE (FUTURE)

PUMPS	
NUMBER	DUTY
PA1	RAW SEWAGE FEED – A SIDE
PB1	RAW SEWAGE FEED – B SIDE
PA2	DENITRIFIED MIXED LIQUOR RECYCLE - A SIDE
PB2	DENITRIFIED MIXED LIQUOR RECYCLE - B SIDE
PA3	RETURN BIOLOGICAL SLUDGE - A SIDE
PB3	RETURN BIOLOGICAL SLUDGE – B SIDE
PA4	FIXED GROWTH REACTOR No.1 (FGR1) RECYCLE - A SIDE
PB4	FIXED GROWTH REACTOR No.1 (FGR1) RECYCLE – B SIDE
PA5	FIXED GROWTH REACTOR No.2 (FGR2) RECYCLE - A SIDE
PB5	FIXED GROWTH REACTOR No.2 (FGR2) RECYCLE – B SIDE
PA6	FIXED GROWTH REACTOR (RX) EFFLUENT – A SIDE
PB6	FIXED GROWTH REACTOR (FGR) EFFLUENT – B SIDE
PA7	CRUDE SLUDGE FEED – A SIDE
PB7	CRUDE SLUDGE FEED – B SIDE
PA8	NITRIFIED MIXED LIQUOR RECYCLE (FUTURE) - A SIDE
PB8	NITRIFIED MIXED LIQUOR RECYCLE (FUTURE) – B SIDE




PA4
PA5
FGRR
FGR1 A SIDE
FGR2 A SIDE
FGR EFFLUENT SUMP A SIDE
PA6
FGR1 B SIDE
FGR2 B SIDE
FGR EFFLUENT SUMP B SIDE
PB6
FGRE
PB4
PB5
ARE
PA2
ANOXIC REACTOR
A SIDE
WEIR
REAERATION REACTOR
A SIDE
FINAL CLARIFIER
WEIR
A SIDE
FCE
PA8
NR
PA3
RBS
WBS
PB3
LABORATORY AREA
FRONT OF TRAILER
PB8
ANOXIC REACTOR
B SIDE
REAERATION REACTOR
B SIDE
FINAL CLARIFIER
B SIDE
PB2

4 METHODOLOGY

4.1 Monitoring of Pilot Plant Performance

The FGR-SGR pilot plant was designed to treat domestic wastewater in two identical process trains, each configured as shown earlier in Figure 2. As described in Chapter 3, the two modules were designed to allow independent controlled changes in process design and operating parameters, such as the sizes of the anaerobic, anoxic, and aeration suspended growth reactors, and the internal recycle flow rates within the process.

Parallel operation of the two pilot-scale FGR-SGR process trains treating a common waste stream allowed valid comparisons between different design and operational modes (*i.e.* hydraulic retention times in the suspended growth reactors, internal recycle flow rates, and the wasting rate of the suspended growth). After both trains were operated in an identical mode for an acclimatization 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. The *t* test for paired comparisons tests whether or not a significant difference exists in the mean values of two sample sets, where a natural pairing of observations exists. For this study, the natural pairings were between the samples taken at the same location and at the same time from the A and B Sides. The *t* test comparisons were carried out at the 0.05 level of significance.

Overall process performance was monitored by the analysis of 24-hour composite samples of process influent and effluent for five-day biochemical oxygen demand (BOD_5), ammonia nitrogen (NH_4^+), nitrate+nitrite (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. Grab samples were also taken periodically from the sampling ports in the FGR cells, in an attempt to assess the phosphorus removal and nitrification rates across the media profile.

The percent phosphorus by dry weight in the process mixed liquor suspended solids was also monitored, to assess the degree of phosphorus storage by the process biomass. 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 calculated is given in Table 2.

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 2 - Summary of Sampling and Analysis Parameters Continued

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**
Filtered Orthophosphate (PO_4) and Nitrite + Nitrate (NO_x)	Primary Clarifier Overflow	2	4
	Anaerobic Zone Effluent	2	4
	Anoxic Zone Effluent	2	4
	FGR Cell 1 Effluent	2	4
	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 Nitrogen (TKN)	Raw Plant Influent	2	4
	Primary Clarifier Overflow	2	4
	Final Clarifier Overflow	2	4
	Total	**6**	**12**
Ammonia Nitrogen (NH_4)	Raw Plant Influent	2	4
	Primary Clarifier Overflow	2	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 1	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 were 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.

4.2 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 conditions in the pilot-scale suspended growth reactors. 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 t=0 in a ratio of 2:1 in the batch reactor, which was then held under fully mixed, anaerobic conditions for a duration which matched the actual hydraulic retention time (HRT) of the anaerobic process 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 mixed liquor from the anaerobic reactor in a ratio of 1:3 (the return biosolids being the flow from the secondary clarifier to the anoxic reactor). For a typical batch test, at the end of the anaerobic phase, an aliquot of return settled biosolids was added to 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 aerobic HRT of the process suspended growth aerobic reactors (*i.e.*, the actual HRT in the FGR sumps and aeration reactor). Again, samples were withdrawn at regular intervals and analyzed for the appropriate parameters.

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.

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. 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.

4.3 Sample Preservation and Analysis

4.3.1 Biochemical Oxygen Demand

Samples for five-day total 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.

4.3.2 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).

4.3.3 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 fiber filters. The solid residue was oven-dried at 104^0C, and then 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.

4.3.4 Total Kjeldahl Nitrogen

Samples for total kjeldahl nitrogen were frozen immediately after collection. Analysis was conducted on a Lachat QuikChem Automated Ion Analyzer according to QuikChem Method No. 10-107-06-2-E.

4.3.5 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 4^0 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.

4.3.6 Nitrate and Nitrite

Samples for nitrate and nitrite 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; this method gives only the total sum of nitrate plus nitrite, and does not distinguish between the two.

4.3.7 Ammonia

Samples for ammonia 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.

4.3.8 Volatile Fatty Acids

Samples for volatile fatty acids were filtered immediately after collection using Whatman No. 4 filter papers, preserved by adding 1 mL of 2% phosphoric acid solution to 10 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.

4.3.9 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.

4.3.10 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.

4.3.11 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.

5 TASK 1 - OPTIMIZATION OF REACTOR HYDRAULIC RETENTION TIMES

5.1 Pilot Plant Performance - Optimization of Anaerobic and Anoxic HRTs

A summary of the pilot plant design and operating parameters during the investigation of optimum hydraulic retention times (HRTs) is given in Table 3. 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 University of B.C. 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.

During the initial acclimatization phase, when both sides were operated in an identical fashion (August, 1992 through February, 1993), batch test results indicated that an approximate 50% reduction in the anoxic volume might be beneficial. Accordingly, the volume of the anoxic basin on the B side process train was halved on March 8, 1993, reducing the actual anoxic 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 acclimatization phase (Table 3).

During Phase 1, batch test investigations indicated that the anaerobic volume in the pilot plant could also be substantially reduced. After the acclimatization 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).

Table 3-Summary of Pilot Plant Design and Operating Parameters - Task 1

Phase	Design Parameters							
Aug/92 - Mar 7/93	Acclimatization - both sides identical - see A Side Phase 1 for parameters							
Phase 1 Mar 8/93 to May 4/93	Reactor	Volume (L)		Nominal HRT (hr)[1]		Actual HRT (min)[2]	
		A Side	B Side	A Side	B Side	A Side	B Side
	Anaerobic	630	630	2.2	2.2	45	45
	Anoxic	1260	630	4.4	2.2	65	35
	Aerobic[3]	1520	1520	5.3	5.3	160	160
	Total	3410	2780	11.9	9.7		
May 5/92 - Jun 5/93	Acclimatization - both sides identical - see B Side Phase 1 for parameters							
Phase 2 Jun 6/93 to Jul 30/93	Reactor	Volume (L)		Nominal HRT (hr)[1]		Actual HRT (min)[2]	
		A Side	B Side	A Side	B Side	A Side	B Side
	Anaerobic	630	380	2.2	1.3	45	25
	Anoxic	630	630	2.2	2.2	35	35
	Aerobic[3]	1520	1520	5.3	5.3	160	160
	Total	2780	2530	9.7	8.8		
Jul 31/93- Nov 29/93	Acclimatization - both sides identical - see B Side Phase 2 for parameters							
	Reactor	Volume (L)		Nominal HRT (hr)[1]		Actual HRT (min)[2]	
Phase 3 Nov 30/93 to Feb 23/94		A Side	B Side	A Side	B Side	A Side	B Side
	Anaerobic	125	380	0.4	1.3	8	25
	Anoxic	630	630	2.2	2.2	35	35
	Aerobic[3]	1520	1520	5.3	5.3	160	160
	Total	2275	2530	7.9	8.8		

[1] Reactor volume/process influent flow rate
[2] Reactor volume/(process influent flow rate +recycle flow rate)
[3] Includes FGR sumps and aeration basin

The removal of total suspended solids (TSS) from the process influent (primary clarifier overflow) throughout the entire study (Phases 1-3 and subsequent investigations) is summarized in Figure 3. As shown, plant effluent (secondary clarifier effluent) TSS concentration was consistently less than 20 mg/L. The performance of the A Side was similar to that of the B Side. The mean process influent and effluent TSS concentrations and percent removals throughout Phases 1-3 are shown in Table 4. The results of the *t* test comparison of the A and B Side means (not shown in Table 4) indicated that the successive reductions in HRT had no significant effect on plant effluent TSS concentration.

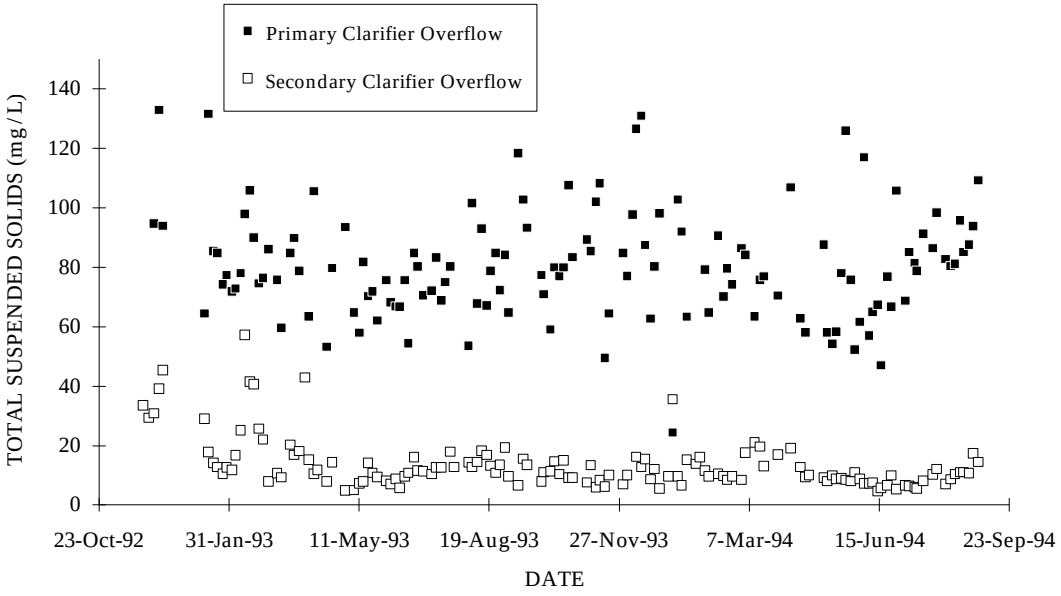


Figure 3 Total Suspended Solids Removal - B Side Process Train

The removal of total five-day biochemical oxygen demand (BOD_5) for the B Side is summarized in Figure 4. Effluent total BOD_5 throughout the study was consistently less than 10 mg/L on both the A and B Sides. The mean process influent and effluent BOD_5 concentrations throughout Phases 1-3 are shown in Table 4. Similar to the results for TSS, the *t* test comparison of the A and B Side means (not shown in Table 4) indicated that the successive reductions in HRT had no significant effect on plant effluent BOD_5 concentration.

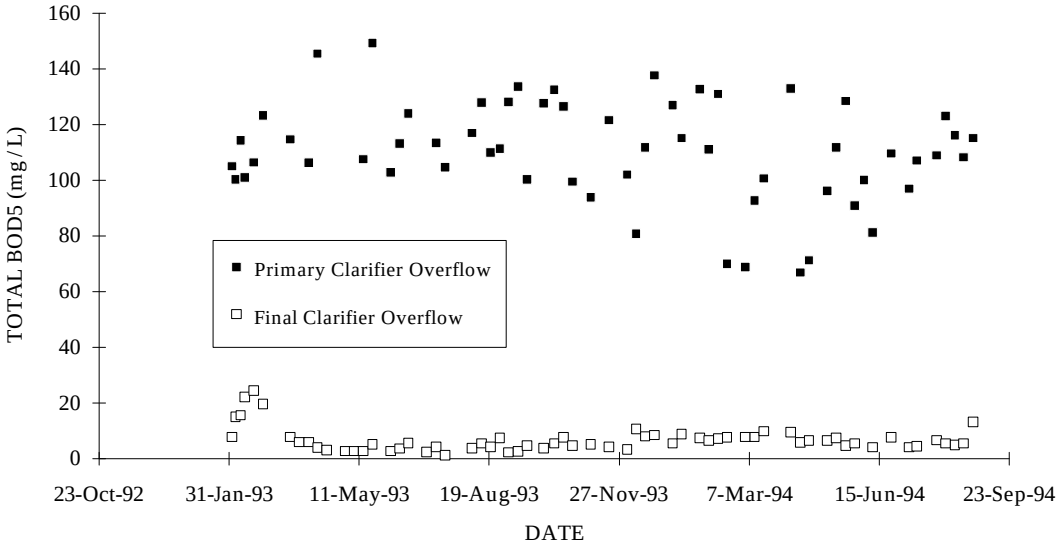


Figure 4 Total BOD_5 Removal - B Side Process Train

Ammonia removal for the B Side is summarized in Figure 5. Effluent ammonia concentration for both the A and B Sides was consistently less than the detection limit of 0.01 mg N / L throughout the study. The process influent and effluent mean ammonia concentrations for both the A and B Sides during Phases 1-3 are shown in Table 4. The *t* test comparison detected no significant difference in ammonia removals between the two sides during Phases 1-3.

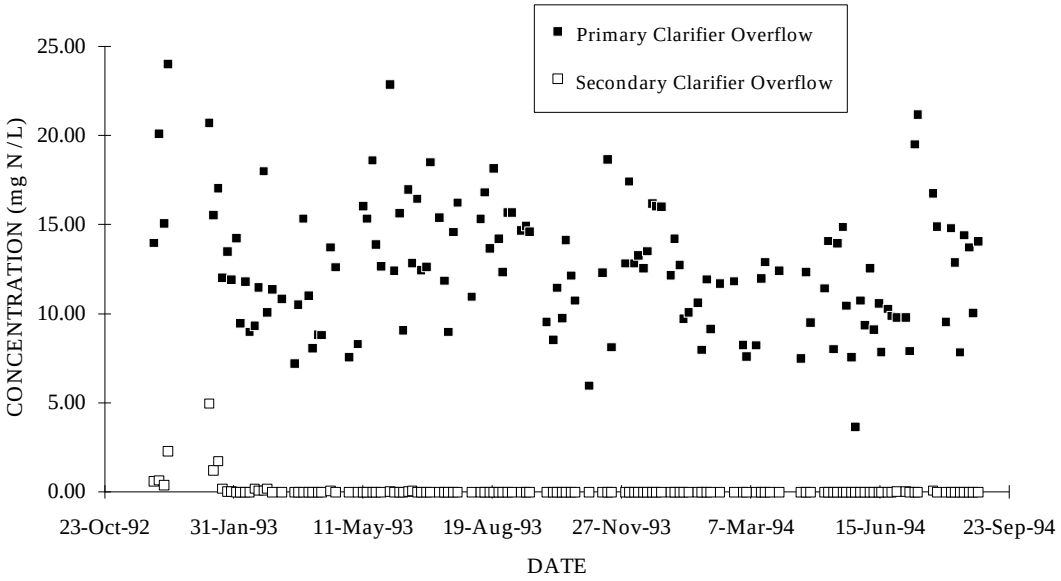


Figure 5 Ammonia Removal - B Side Process Train

Removal of total kjeldahl nitrogen (TKN) for the B Side process is shown in Figure 6. As shown, effluent TKN concentration (filtered samples) was consistently less than 5 mg N/L. Similar results were recorded for the A Side. The mean influent and effluent TKN concentrations for the A and B Sides throughout Phases 1-3 are compared in Table 4. The *t* test results indicated that the reductions in HRT had no significant effect on TKN removal.

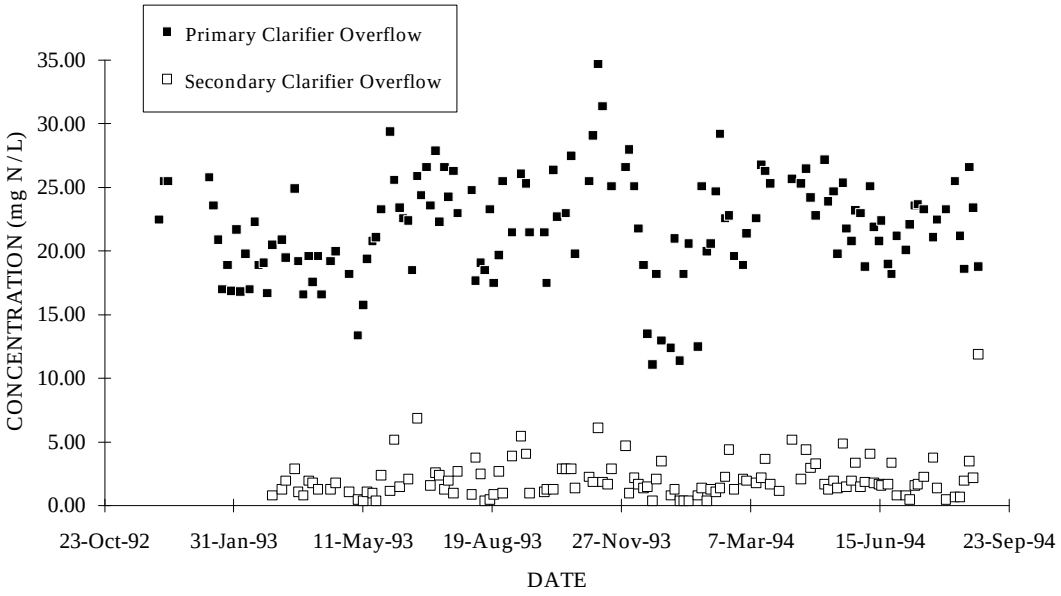


Figure 6 Total Kjeldahl Nitrogen Removal - B Side Process Train

Process removal of total phosphorus and plant effluent orthophosphate concentration are summarized in Table 4. Mean percent removal of total phosphorus was highest during Phase 3 (91% for the A Side and 94% for the B Side). Similarly, plant effluent mean orthophosphate concentration was lowest during Phase 3 (0.27 mg P/L and 0.10 mg P/L for the A and B Sides, respectively). The results of the *t* test did not show a significant difference between the A and B Side phosphorus removal during Phase 3. Since the lowest effluent orthophosphate concentrations were observed during Phase 3, it appears that the progressive reductions in HRT had a beneficial effect on phosphorus removal; this conclusion is supported to some extent by the batch test data presented later in this chapter. However, the high degree of phosphorus removal observed during Phase 3 is mainly attributed to a factor unrelated to the suspended growth HRT, as described below.

Phosphorus removal for the A and B Side process trains, including Phases 1, 2, and 3, and subsequent periods, is summarized in Figures 7a and 7b, respectively. Some of the short-term increases in effluent orthophosphate

shown in Figure 7 can be attributed to equipment failures. However, following a period of good to excellent 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 1 mg P/L (see Figure 7). The increase in effluent orthophosphate was not associated with equipment failures, and effluent parameters other than phosphorus did not increase.

Table 4 - Summary of Pilot Plant Performance During Task 1

Parameter	**Sample**	**Mean Concentration (24 hr composite sample)**					
		Phase 1		Phase 2		Phase 3	
		A Side	B Side	A Side	B Side	A Side	B Side
Total Suspended Solids (mg/L)	Influent[1]	84	101	71	90	85	83
	Effluent[2]	24	15	14	12	12	11
	% Removal	71	85	80	87	86	87
Total BOD_5 (mg/L)	Influent[1]	97	96	104	105	112	111
	Effluent[2]	8	5	5	4	8	7
	% Removal	92	95	96	97	93	94
Ammonia (mg N/L)	Influent[1]	11.0	11.0	14.5	14.0	12.5	12.6
	Effluent[2]	<0.01	<0.01	<0.01	0.01	<0.01	<0.01
	% Removal	>99	>99	>99	>99	>99	>99
Tot. Kjeldahl Nitrogen (mg N/L)	Influent[1]	19.0	19.3	23.7	24.1	21.5	21.2
	Effluent[2,3]	1.5	1.6	2.8	2.4	2.5	1.8
	% Removal	87	90	78	87	91	90
Total Phosphorus (mg P/L)	Influent[1]	3.3	3.3	3.8	3.9	4.0	3.9
	Effluent[2,3]	1.35	0.95	0.88	0.12	0.35	0.25
	% Removal	59	71	77	97	91	94
Orthophosphate (mg P/L)	Effluent[2]	0.62	0.38	0.53	0.17	0.27	0.10

[1] Process influent is primary clarifier overflow
[2] Plant effluent is secondary clarifier overflow
[3] Filtered Sample

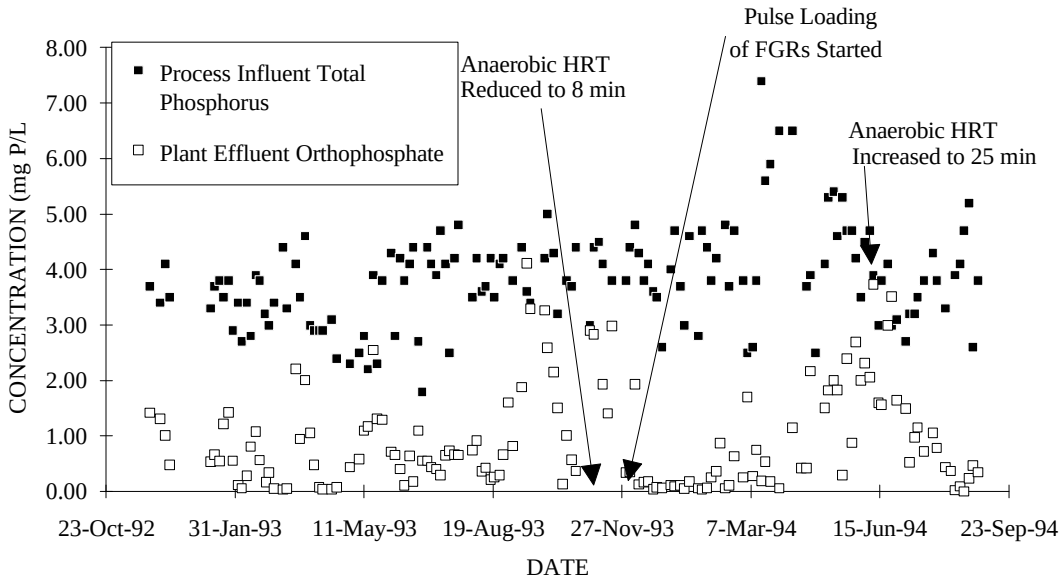


Figure 7a - Pilot Plant Phosphorus Removal - A Side Process Train

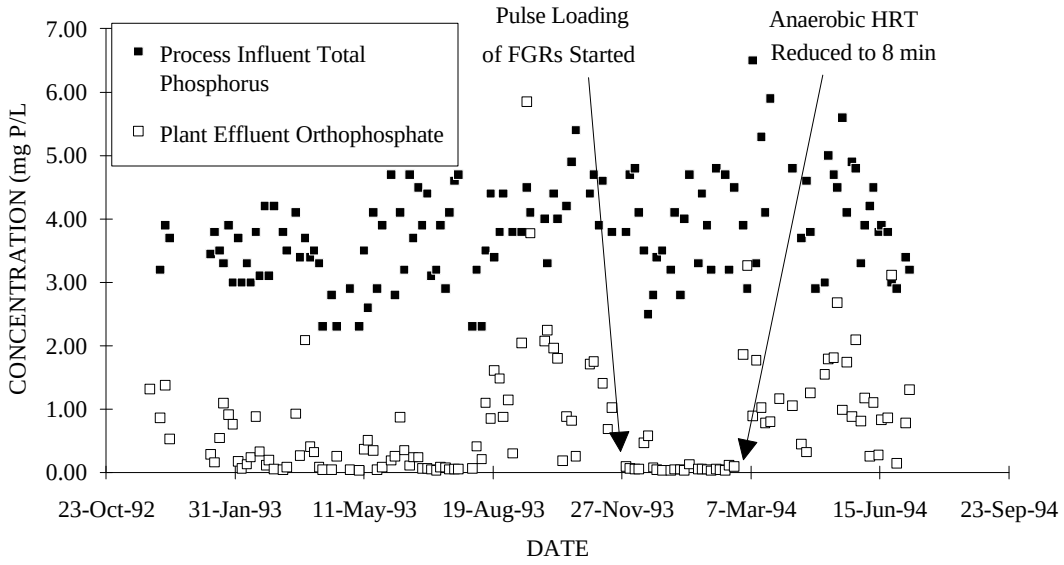


Figure 7b - Pilot Plant Phosphorus Removal - B Side Process Train

Investigations into the cause of the deterioration in biological phosphorus removal revealed an excessive buildup of biosolids on the FGR media, probably caused by suspended organisms in the mixed liquor adhering to the biofilm on the media. Accordingly, a hydraulic pulse loading regime was introduced on both the A and B Sides of the pilot plant, whereby the FGR irrigation (recycle) rate was periodically increased for a short period each day, to flush accumulated solids from the media. Regular pulse loading of the

FGRs beginning in late November of 1993 resulted in an immediate and consistent improvement in biological phosphorus removal in both process trains, to effluent mean orthophosphate concentrations of 0.27 mg P/L on the A Side and 0.10 mg P/L on the B Side during Phase 3 (Table 4 and Figure 7).

During Phases 1 and 2 and during the acclimatization periods before Phases 2 and 3, the process operating MLSS concentration was subject to wide fluctuations, despite a constant daily sludge wasting rate. In addition to improving biological phosphorus removal, pulse loading of the FGRs also reduced the fluctuations in process MLSS concentration. Further, infestations of filter flies (*psychoda*) on the FGR media during warm weather were greatly reduced after the pulse loading regime was introduced.

For the FGR-SGR process, excessive suspended solids buildup on the media could lead to anaerobic conditions deep within the FGR biofilm, causing phosphate release without associated carbon storage by suspended bacteria which adhere to the biofilm, and then become buried as solids buildup progresses. Alternatively, suspended bacteria trapped on the FGR biofilm for relatively long periods could release phosphate under aerobic conditions; it has been reported by others that extended periods of aeration in biological phosphorus removal systems can lead to aerobic phosphate release (Comeau *et. al.*, 1987).

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 (Table 4 and Figure 7a). However, following the reduction in anaerobic HRT on the B Side, an immediate deterioration in phosphorus removal was observed (Figure 7b). At a slightly earlier date, phosphorus removal also began to deteriorate on the A Side (Figure 7a). Therefore, it appeared that a change in operating condition had occurred near the end of Phase 3 which rendered the 8 minute anaerobic HRT insufficient for good phosphorus removal. A possible explanation for the deterioration in phosphorus removal is given below.

Biological phosphorus removal processes depend on the accumulation of phosphorus by process bacteria in amounts greater than that required for reproduction and growth. As described in Chapter 3, the phosphorus accumulation mechanism is believed to depend on the uptake and storage of easily-degradable carbon in the anaerobic phase by the process bacteria, the uptake and storage of carbon being accompanied by the degradation of intracellular phosphorus reserves and the release of orthophosphate to the bulk solution. In the subsequent aerobic phase, the stored carbon is oxidized and part of the resulting energy is used to build phosphate reserves within the cell.

The degree of bacterial phosphate storage in the aerobic phase is believed to depend in part upon the degree of bacterial carbon storage in the anaerobic phase. That is, increasing the availability of soluble carbon in the anaerobic phase should increase the phosphorus removal capability of the process bacteria in the subsequent aerobic phase.

The concentration of total volatile fatty acids (VFA) in the influent (primary clarifier overflow) to the B Side process train is shown in Figure 8. Also shown in Figure 8 is the liquid temperature of the primary sludge fermenter. The pattern on the A Side was similar to that of the B Side.

As shown in Figure 8, the VFA concentration in the process influent was generally in the range 30-50 mg/L (as acetic acid) during the period from May of 1993 to February of 1994. During this period, the fermenter temperature was mainly in the range 15-20 degrees Celcius, and phosphorus removal was effective on both the A and B Sides, except for the period from July-November of 1993, when excess accumulation of solids on the FGR media was observed (see previous discussion and Figure 7).

In February of 1994, the fermenter temperature began to fall into the range 10-15 degrees Celcius, and the VFA concentration in the process influent fell into the range 15-35 mg/L (Figure 8). It is likely that the lower fermenter temperature led to a lower biochemical reaction rate in the primary sludge fermenters, reducing the amount of VFA added to the primary clarifier overflow by the fermenter loop. At around the same time, phosphorus removal on the A Side (which was being operated with an 8 minute anaerobic HRT) began to deteriorate (Figure 7a). Phosphorus removal on the B Side (which was being operated with a 25 minute anaerobic HRT) appeared to be unaffected. However, as soon as the anaerobic HRT on the B Side was reduced to 8 minutes, phosphorus removal began to deteriorate (Figure 7b).

In an effort to improve phosphorus removal by allowing a greater opportunity for anaerobic carbon storage, the actual anaerobic HRTs on both the A and B Sides were increased from 8 minutes to 25 minutes in June of 1994. At around the same time, the fermenter temperature and the VFA concentration in the process influent began to increase (Figure 8), and a steady improvement in phosphorus removal was observed (Figure 7). It is unclear whether the improvement was due to the increase in anaerobic HRT or to the increase in process influent VFA concentration, or to a combination of the two. The need to complete Tasks 2 and 3 of the study did not allow sufficient time for further investigation of the relative effects of anaerobic HRT and VFA concentration. However, it appears that the process can operate effectively with a very low anaerobic HRT, provided that a sufficiently high concentration of VFA is available in the process influent.

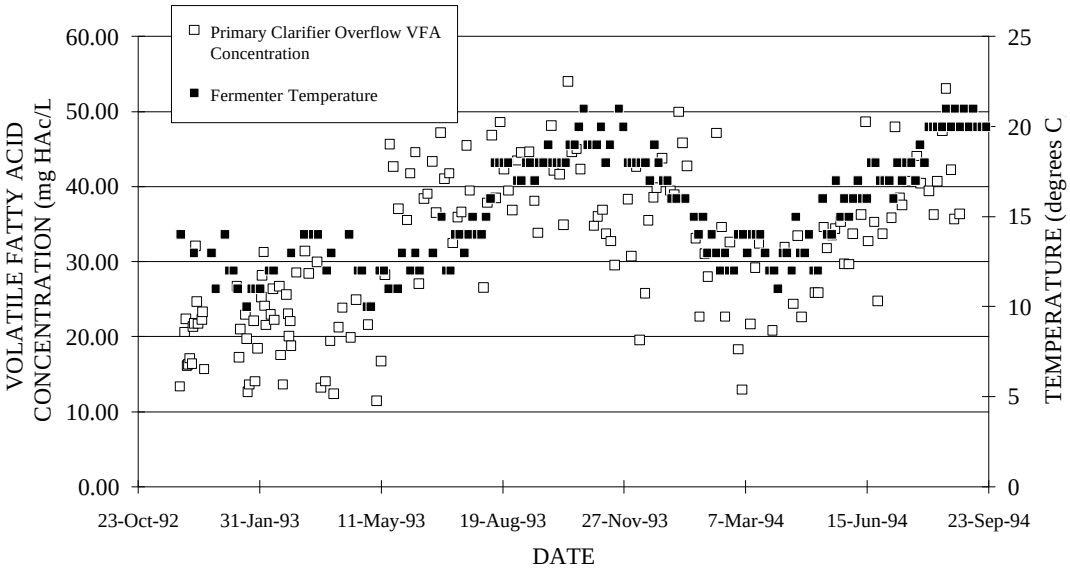


Figure 8 - Volatile Fatty Acid Concentration and Fermenter Temperature in B Side Process Train

5.2 Batch Tests

5.2.1 Optimization of Anoxic Reactor

During the HRT optimization, bench-scale batch tests designed to simulate the performance of the suspended growth process reactors were periodically conducted, to evaluate biochemical reaction rates. Details of the batch test methodology were described earlier (see Chapter 4).

As shown earlier in Table 3, Phase 1 operation compared an actual anoxic HRT of 65 minutes on the A Side to 35 minutes on the B Side. The results of a typical batch test conducted during Phase 1 are shown in Figures 9a (A Side simulation) and 9b (B Side simulation).

In the anoxic phase on the A Side, denitrification of nitrate was accompanied by phosphorus uptake, denitrification-phosphorus uptake being complete within the first 30 minutes, with orthophosphate release observed over the remaining 35 minutes of the anoxic phase (Figure 9a). This implies that the bacteria responsible for enhanced biological phosphorus removal are 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 was complete 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 9b). Anaerobic release of phosphorus without associated carbon uptake and storage has been termed "secondary release," and is postulated to be detrimental to biological phosphorus removal, because there is no stored carbon to drive subsequent uptake of the released phosphorus (Barnard, 1983).

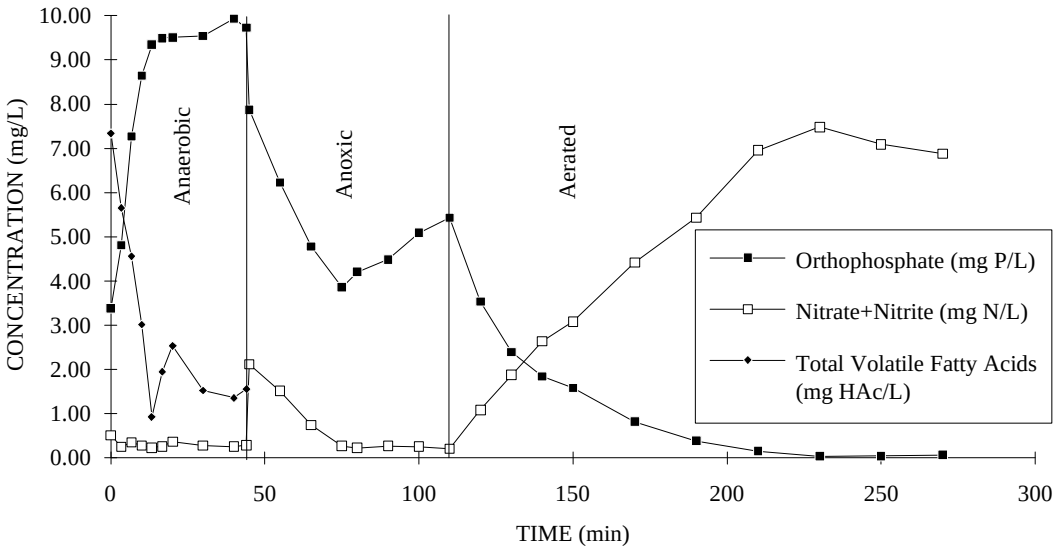


Figure 9a - Phase 1 Batch Test - April 29, 1993 - A Side Process Train

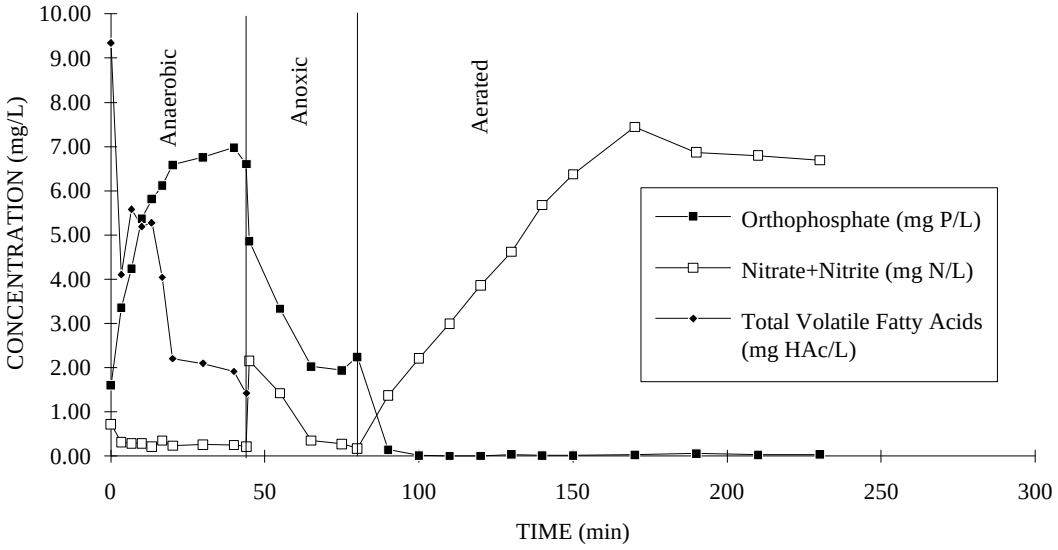


Figure 9b - Phase 1 Batch Test - April 29, 1993 - B Side Process Train

The beneficial effect of reducing the anoxic retention time is clearly shown in Figure 9. For the A side, 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 (*i.e.*, at t=110 minutes, see Figure 9a). 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, however, where considerably less anoxic phosphorus release occurred, and the orthophosphate concentration at the outset of the aerobic phase (*i.e.*, at t=80 minutes, see Figure 9b) was only 2.3 mg P/L, only 10 minutes of aeration was required to reach an orthophosphate concentration of less than 0.5 mg P/L. Note that complete nitrification of ammonia in the aerobic phase by suspended organisms on the B Side took much longer (80 minutes) than phosphorus uptake (10 minutes).

As shown in Figure 9, the rate of phosphorus uptake-storage by the process bacteria using nitrates as an electron acceptor in the anoxic phase was similar to the initial, rapid rate of phosphorus uptake using dissolved oxygen as an electron acceptor in the aerobic phase. This implies that aeration requirements might be reduced in full-scale BNR plants which include nitrification, by maximizing the use of any nitrates produced for phosphorus uptake, thereby minimizing the requirement for dissolved oxygen.

5.2.2 Optimization of Anaerobic Reactor

As shown in Table 4 and Figure 9, the behavior in the anaerobic zone during Phase 1 was similar for both the A side (Figure 9a) and the B side (Figure 9b). That is, anaerobic release of phosphorus appeared to occur at two distinct rates. Most of the VFA uptake with associated (rapid) phosphorus release was complete within the first 20 minutes for both A and B sides; after that point, VFA uptake-phosphorus release continued at a much slower rate. The phosphorus release observed after the completion of denitrification might represent secondary release (*i.e.*, without associated carbon storage). Alternatively, the slower rate of anaerobic phosphorus release might be an indication of the rate of production of VFA by fermentation in the anaerobic reactor, with the fermented VFA being taken up and stored with associated phosphorus release as soon as it became available. In any case, there appeared to be little benefit in extending the anaerobic phase beyond the initial, rapid rate of phosphorus release-VFA uptake.

The results of Phase 1 operation suggested that the anaerobic reactor could be substantially reduced, and this was undertaken during Phase 2, where the actual anaerobic HRT was reduced from 45 minutes to 25 minutes on the B Side. The results of batch test simulations (not shown) and process monitoring (Table 4) conducted during Phase 2 indicated that the reduction in anaerobic actual HRT from 45 minutes to 25 minutes did not compromise biological phosphorus removal. Accordingly, the anaerobic volume on the A Side was reduced to give an actual HRT of 25 minutes, and both sides were operated identically for an acclimatization period. For Phase 3, the anaerobic volume on

the A Side was reduced to give an actual HRT of 8 minutes. The results of a typical batch test conducted during Phase 3 to compare the two process trains are shown in Figure 10.

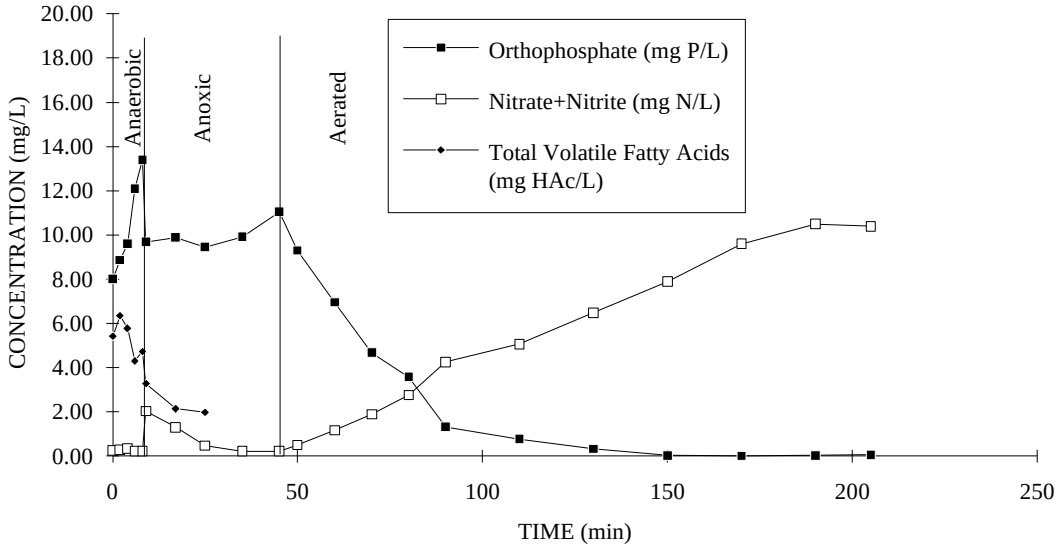


Figure 10a - Phase 3 Batch Test - Dec 2, 1993 - A Side Process Train

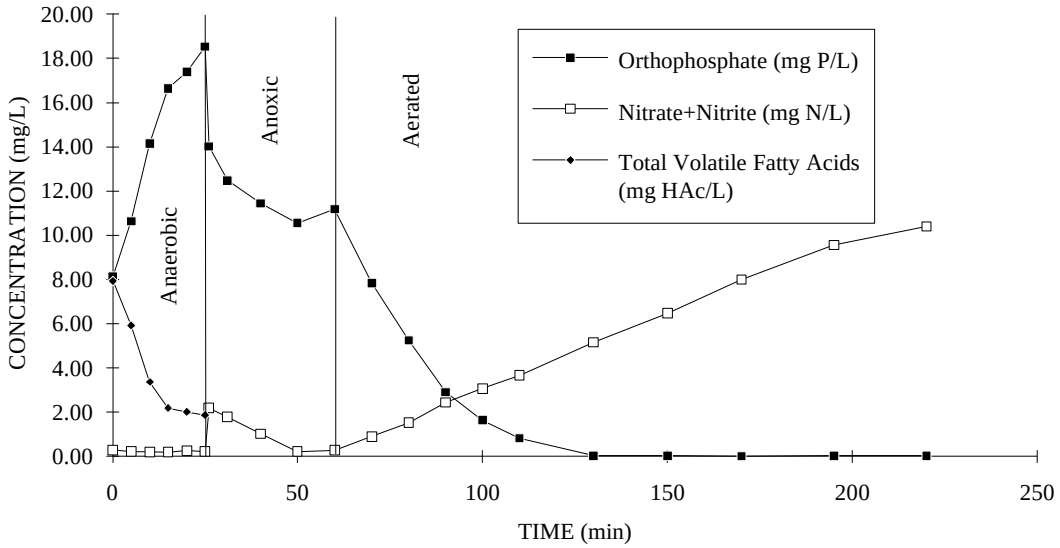


Figure 10b - Phase 3 Batch Test - Dec 2, 1993 - B Side Process Train

The results the of Phase 3 batch tests indicated that the reduction in anaerobic HRT from 25 minutes to 8 minutes on the A Side did not lead to an increase in effluent phosphorus concentration (Table 4 and Figure 10). On the A Side, it appears that the anaerobic HRT of 8 minutes was completed before all of the immediately available VFA had been taken up, since VFA uptake in the subsequent anoxic phase was observed (Figure 10a). On the B Side, VFA uptake with associated rapid phosphorus release was virtually complete after 15 minutes (Figure 10b), and the rate of phosphorus release slowed thereafter, similar to the results shown in Figure 9 for the Phase 1 batch test.

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 (10 mg P/L - Figure 10a) compared to the B Side (14 mg P/L - Figure 10b). 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 10b), similar to the results observed during the Phase I batch test shown earlier in Figure 9. On the other hand, denitrification on the A Side was not accompanied by a net phosphorus uptake (Figure 10a); it may be that the VFA still available in solution at the outset of the anoxic phase on the A Side resulted in a continuation of VFA uptake 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, resulting in little or no net phosphorus uptake in the anoxic zone (this scenario implies the presence of both denitrifying and non-denitrifying bacterial populations capable of excess biological phosphorus removal). The results of mass balances carried out on the anoxic reactor during Phase 3 confirmed that the shorter anaerobic HRT led to reduced anoxic phosphorus uptake. Mean anoxic orthophosphate removal in the pilot plant during Phase 3 was only 3.6 mg P/L on the A Side (8 minute anaerobic HRT), compared to 14.2 mg P/L on the B Side (25 minute anaerobic HRT).

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) 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. It therefore appears that the shorter anaerobic HRT on the A Side (8 minutes) had neither a detrimental nor a beneficial effect on process performance, compared to 25 minutes 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. However, as discussed earlier in this Chapter (Section 5.1), good phosphorus removal with an 8 minute anaerobic HRT appeared to depend on an adequate concentration of VFA in the process influent.

The results shown in Figures 9 and 10 indicate that the aerated actual HRT could be substantially reduced from 160 minutes to 80-90 minutes, without

compromising biological phosphorus removal. The results also indicate that the time required for complete nitrification of ammonia by the suspended biomass was substantially longer than that required for complete phosphorus removal. However, in the FGR-SGR pilot plant, nitrification was mainly accomplished by the fixed growth in the FGRs upstream of the aeration basin. Therefore, the aeration basin HRT could be optimized for phosphorus removal only. Due to project time constraints, the aerated HRT was not further investigated.

5.2.3 Biochemical Reaction Rates Across the FGR Profile

Sampling across the FGR profile 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 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 elsewhere in this report.

6 TASK 2 - OPTIMIZATION OF RECYCLE FLOW RATES

6.1 FGR Recycle

As described in Chapter 5, it was discovered during the investigation to optimize reactor hydraulic retention times 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 (Tasks 2 and 3), regular pulse loading of the FGRs was carried out on both the A and B Sides.

On February 24, 1994, 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 HRT was selected on the basis of the previous work to optimize reactor size (Chapter 5). Both sides were then operated identically for an acclimatization period until April 28, 1994, according to the parameters shown for the A Side during Phase 3 in Table 3. As described in Chapter 5, phosphorus removal on both the A and B Sides deteriorated during the acclimatization period following Phase 3 of Task 1.

However, due to time constraints, and since the primary purpose of the next step in the investigation was to determine the effect of FGR irrigation rate on nitrification of ammonia, and since the process upset of phosphorus removal did not affect ammonia removal (see Figure 5 in Chapter 5), the investigation to optimize the FGR recycle rate was carried forward, despite the phosphorus removal upset.

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), to investigate the effect of a lower FGR recycle rate on removal of biochemical oxygen demand, nitrification of ammonia, and biological phosphorus removal. All other parameters were maintained equal between the two sides.

The performance of the A and B Side FGRs (trickling filters) during the comparison of recycle flow rates (Phase 4) is summarized in Table 5. The mean removals across each of FGR Cells 1 and 2 of filtered BOD_5, ammonia, and orthophosphate were compared between the A and B Sides using the *t* test, to determine if a statistically significant difference existed. Similarly, the

removals of total BOD_5 and total phosphorus across the entire process were compared between the A and B Sides.

The results of the *t* test (not included in Table 5) indicated that the lower FGR recycle rate on the B Side FGR did not result in a significant reduction in BOD removal effectiveness. Mean removals of filtered BOD_5 were 5-6 mg/L across each FGR cell; removal of total BOD_5 across the entire process was 93 mg/L for the A Side, and 97 mg/L for the B Side.

Mean removal of ammonia across FGR Cell 1 was 7.1 mg N/L for the A Side, and 5.8 mg N/L for the B Side (Table 5). For FGR Cell 2, the mean ammonia removal rates were 3.2 mg N/L for both the A and B Sides. Ammonia removal across the entire process was 11.0 mg N/L for the A Side and 12.2 mg N/L for the B Side. According to the results of the *t* test, none of the above comparisons showed a significant difference in ammonia removal between the A and B Sides.

Table 5 - Summary of FGR Performance During Task 2

Parameter	Mean Removal or Production (mg/L)					
	FGR Cell 1		FGR Cell 2		Total Process	
	A Side	B Side	A Side	B Side	A Side	B Side
Filtered BOD_5 Removal	6	5	6	6	N/A[1]	N/A[1]
Total BOD_5 Removal	N/A[1]	N/A[1]	N/A[1]	N/A[1]	93	97
Ammonia Removal (as N)	7.1	5.8	3.2	3.2	11.0	12.2
Nitrate+Nitrite Production (as N)	11.5	9.7	3.8	4.9	7.0	7.4
Orthophosphate Uptake (as P)	5.1	5.6	2.0	3.1	23.3	22.5
Total Phosphorus Removal (as P)	N/A[1]	N/A[1]	N/A[1]	N/A[1]	2.0	2.6

[1] Not Applicable

According to the *t* test, the mean mass of orthophosphate uptake across the A Side FGR Cell 1 (5.1 mg P/L) was not significantly different than that of the B Side Cell 1 (5.6 mg P/L). However, the mean orthophosphate uptake across the A Side FGR Cell 2 (2.0 mg P/L) was significantly lower than that of the B Side Cell 2 (3.1 mg P/L). The removal of total phosphorus across the entire

process was not significantly different between the A Side (2.0 mg P/L) and the B Side (2.6 mg P/L).

It appears that the reduction in FGR recycle rate on the B Side did not result in a deterioration in process removals of BOD, ammonia, or phosphorus. Although the mean ammonia removal was slightly lower across the B Side FGR Cell 1 than on the A Side (Table 5), the difference was not found to be statistically significant, possibly due to the relatively wide variance in the data. Removal of ammonia across the entire process was similar on both sides. In a process heavily loaded with ammonia, the lower nitrification rate at the lower FGR Cell 1 recycle rate might be of concern. However, in the case of the pilot plant, the ammonia concentration in the process influent was consistently nitrified to less than 1 mg N/L of ammonia across FGR Cells 1 and 2, and to non-detectable amounts across the aeration basin, indicating an excess of nitrification capacity. Further, phosphorus removal was higher across the B Side FGRs than on the A Side.

In light of the above results, it was concluded that the lower FGR recycle rate could be used without compromising process performance. Since a lower pumping rate results in lower operating costs, it makes economical sense to maintain the lowest flow rates that do not interfere with process performance. An FGR irrigation rate even lower than the 14 L/min used on the B Side would result in incomplete wetting of the media, which would also be uneconomical. The lower recycle rate of 14 L/min used on the B Side during Phase 4 was therefore determined to be the optimum.

6.2 Denitrified Recycle Rate

In the FGR-SGR pilot plant, the denitrified recycle was the internal process flow which returned denitrified mixed liquor from the anoxic reactor to the anaerobic reactor (see Figure 2). Time did not allow operation of the pilot plant at several different denitrified recycle flow rates, with time for acclimatization and steady-state monitoring after each change. However, a batch test simulation was developed to investigate at bench-scale the short-term effects of varying the denitrified recycle rate.

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 hydraulic retention time in the process bioreactor(s), since average actual hydraulic retention time 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).

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 2). Besides reducing the actual HRT, an increase in process influent flow rate also tends to dilute the MLSS concentration in the process reactor, increasing the food to microorganism (F/M) ratio. The batch test in this case was designed to investigate the short-term 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, thereby reducing the F/M ratio.

At the time of the batch test (August 23, 1994), both the A and B Side process trains were being operated according to the parameters described for the B Side during Phase 3 (see Table 3). That is, the process influent flow (Q) was 4.8 L/min, the denitrified recycle rate was 9.6 L/min (2Q), and the actual anaerobic retention time was 25 minutes. As described later in Chapter 7, the B Side process was being operated at a lower MLSS concentration than the A Side during the period when the batch test was carried out.

For the batch test, five batch reactors were operated in parallel according to the procedure described earlier in Chapter 4. The batch test parameters are summarized in Table 6. The five tests were conducted simultaneously, using grab samples taken from the A Side process train only. The designated aliquot of sample for each reactor was taken from a common, fully mixed grab sample of process liquid from the A Side.

Table 6 - Summary of Parameters for Investigation of Denitrified Recycle Rate

Reactor #	Simulated Denitrified Recycle (times Q[1])	Volume of Sample Added (mL)			Actual Hydraulic Retention Time (min)		
		Process Influent	Anoxic Liquor	Return Biosolids	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

The simulated daily peak load for the batch test was assumed to be equal to 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 operated 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 operated to simulate the effects of increasing the denitrified recycle rate to 2.5Q and 3Q, respectively, to decrease the dilution of MLSS. The simulated process influent flow rate was 1.65Q for all five batch reactors.

The results of the batch test investigation of denitrified recycle rate are summarized in Figures 11a-11e. For the control reactor (#1-Figure 11a), the simulated increase in process influent from Q to 1.65 Q did not appear to result in a deterioration in system performance. The actual anaerobic HRT of 22 minutes was just long enough for complete removal of all volatile fatty acids (VFA) from solution, with associated orthophosphate release. The actual anoxic HRT of 28 minutes was more than adequate for complete denitrification of all available nitrate+nitrite, with associated orthophosphate uptake. The actual aerated 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 had enough reserve capacity to handle a daily peak flow of 1.65Q, at least over the short term.

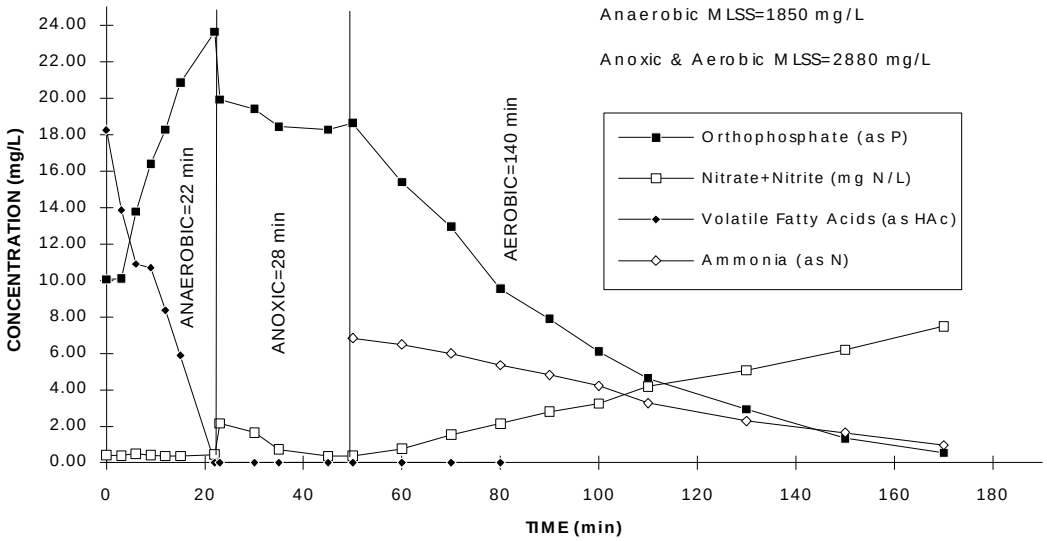


Figure 11a - Batch Reactor #1-Simulated Denitrified Recycle=2Q

The simulated effects of reducing the denitrified recycle from 2Q to 1.5Q (Reactor #2) are shown in Figure 11b. The decrease in recycle resulted in an increase in anaerobic actual HRT from 22 minutes to 25 minutes. However, the decrease in recycle also resulted in a decrease in anaerobic MLSS concentration to 1680 mg/L (Figure 11b), compared to 1850 mg/L in the control reactor (Figure 11a). 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 Chapter 3, greater anaerobic storage of carbon (*e.g.*, VFA) by bacteria should result in greater 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 11b (*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.

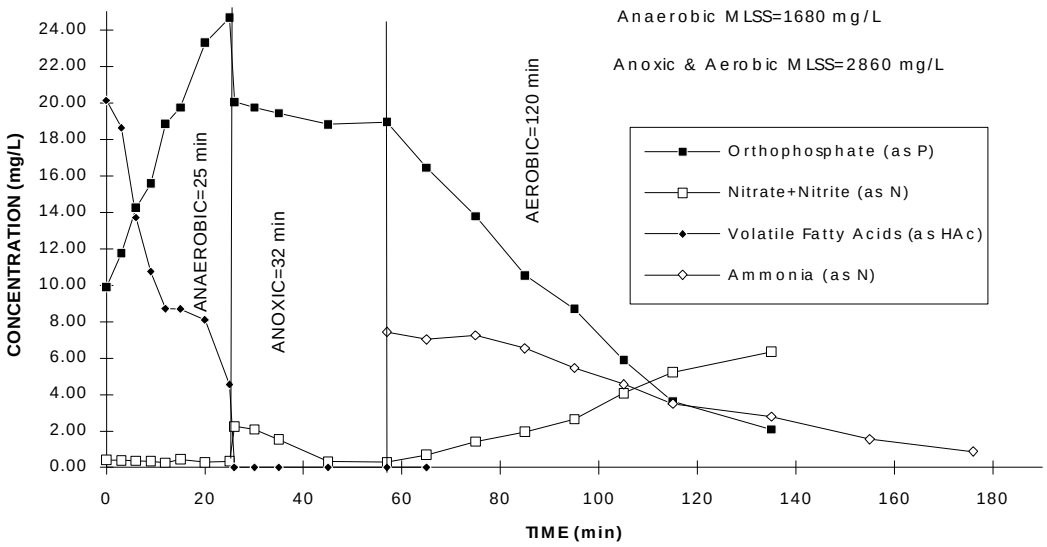


Figure 11b-Batch Reactor #2-Simulated Denitrified Recycle=1.5Q

The behavior in the anoxic phase for Reactor #2 was similar to that observed in Reactor #1; all available nitrate+nitrite was removed, with associated orthophosphate uptake. 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 11b), indicating similar phosphorus removal effectiveness to Reactor #1 in the short term.

The simulated effects of reducing the denitrified recycle to Q (Reactor #3) are shown in Figure 11c. In this case, the decrease in recycle increased the anaerobic HRT to 30 minutes, but reduced the anaerobic MLSS concentration to 1320 mg/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, phosphate release in Reactor #3 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 11c). After the VFA concentration in the anoxic phase reached zero, phosphate uptake was observed until all available nitrates were denitrified. At the end of the aerated phase, the bulk solution orthophosphate concentration was 1.7 mg P/L, which was significantly higher than in Reactor #1.

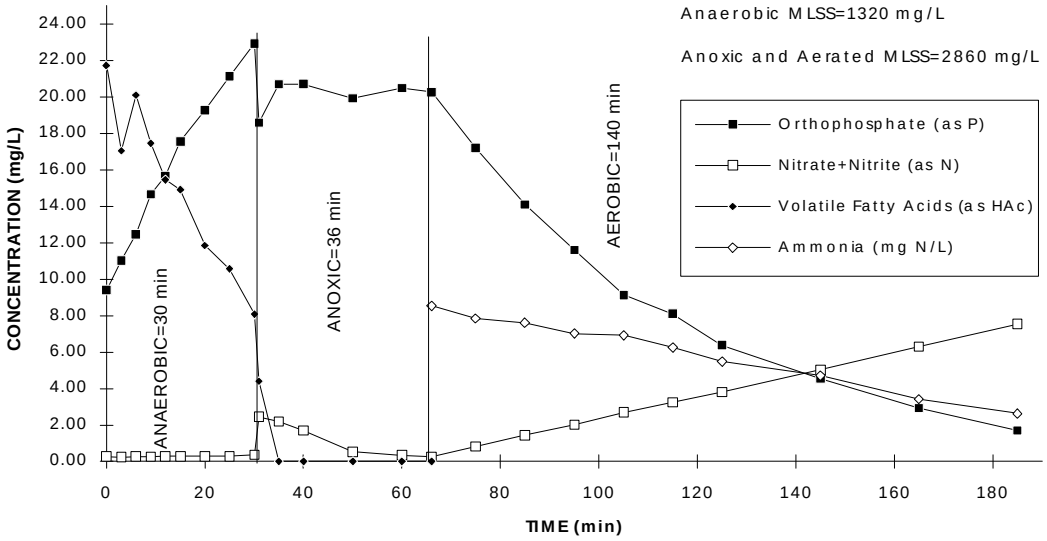


Figure 11c -Batch Reactor #3 - Simulated Denitrified Recycle=Q

The effects of increasing the denitrified recycle to 2.5Q (Reactor #4) are shown in Figure 11d. In this case, the anaerobic MLSS concentration was increased to 2150 mg/L (Reactor #4-Figure 11d), compared to 1850 mg/L in the control reactor (#1 - Figure 11a), and the actual anaerobic HRT was reduced to 19 minutes, compared to 22 minutes in the control. As shown in Figure 11d, 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 11d). 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 (#3).

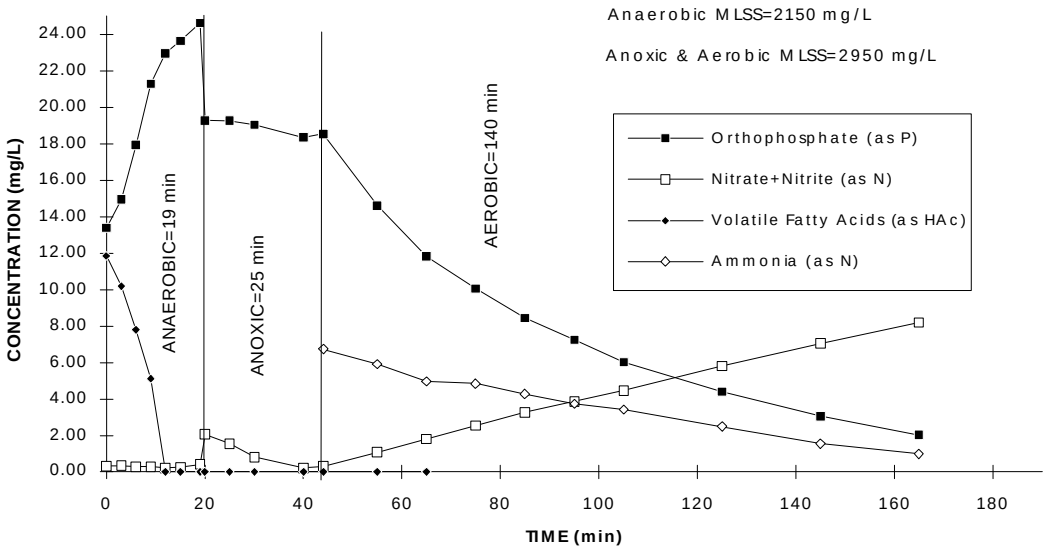


Figure 11d-Batch Reactor #4-Simulated Denitrified Recyle=2.5Q

The simulated effects of further increasing the denitrified recycle to 3Q (Reactor #5) are shown in Figure 11e. The sample for anaerobic MLSS for Reactor #5 was lost. However, the value was estimated 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. 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 actual anoxic 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.

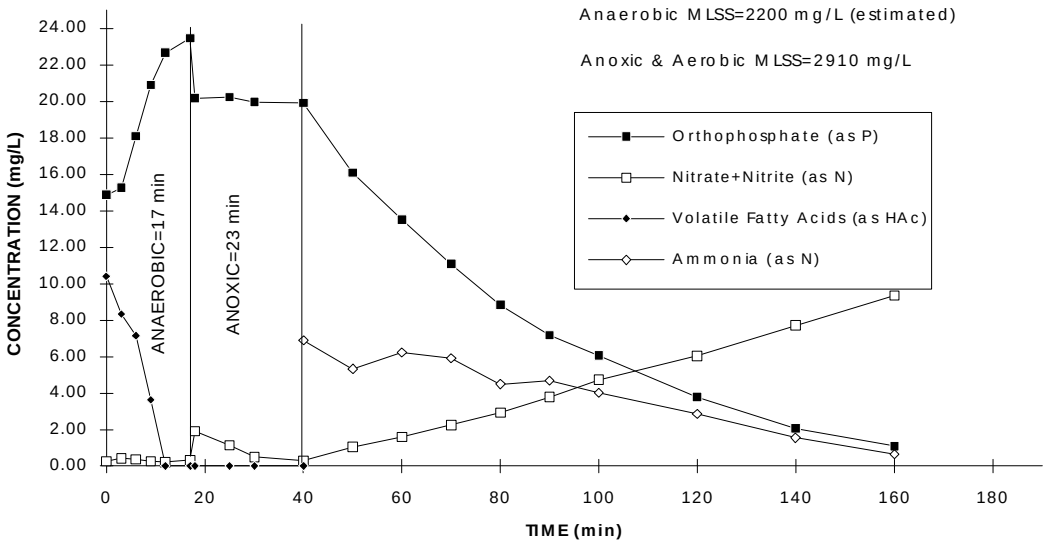


Figure 11e-Batch Reactor #5-Simulated Denitrified Recycle=3Q

From the above results, it appears that reducing the denitrified recycle rate to buffer the effects of hydraulic shocks caused by increases in process influent flow rate would not be an effective maneuver. The increase in anaerobic HRT gained by reducing the denitrified recycle appeared to be more than offset by a decrease in anaerobic MLSS 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 phosphate loading to the aerobic phase.

On the other hand, increasing the denitrified recycle to reduce the dilution of the anaerobic MLSS caused by an increase in process influent flow rate resulted in a higher degree of bacterial VFA uptake-storage. The result was faster removal of available VFA during the anaerobic phase. As discussed in Chapter 3, 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 the longer term.

If the batch tests described above provided a reasonably accurate simulation of the short-term effects of manipulating the denitrified recycle rate, it appears 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. It remains to be seen whether this would result in improved phosphorus removal over the longer term; however, current theory suggests that this should be the case.

6.3 Settled Biosolids Recycle Rate

The settled biosolids recycle in the FGR-SGR pilot plant was the internal process flow which returned settled biosolids from the final clarifier to the anoxic reactor (see Figure 2). 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 acclimatization and steady-state monitoring after each change. However, a batch test simulation was developed to investigate at bench-scale the short-term effects of varying the biosolids recycle rate.

In the FGR-SGR pilot plant, the actual HRT in the 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 2). 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 HRTs, and also to investigate the effects of increasing the biosolids recycle to dampen the dilution effects of the 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 29, 1994), both the A and B Side process trains were being operated according to the parameters described for the B Side during Phase 3 (see 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), and the settled biosolids recycle rate was 4.8 L/min (Q); the actual anoxic HRT was 35 minutes, and the actual aerated HRT was 160 minutes.

For the batch test, five batch reactors were operated in parallel according to the procedure described earlier in Chapter 4. The batch test parameters are summarized in Table 7. All aliquots of process liquid for the batch test were taken from common, completely mixed samples taken from the A Side only.

Table 7 - Summary of Parameters for Investigation of Settled Biosolids Recycle Rate

Reactor #	Simulated Settled Biosolids Recycle (times Q[1])	Volume of Sample Added (mL)			Actual Hydraulic Retention Time (min)		
		Process Influent	Anoxic Liquor	Return Biosolids	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

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 daily peak load was again assumed to be equal to 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 anoxic HRT from 35 minutes to 28 minutes, and a reduction in actual aerated 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.

The results of the batch test investigation of settled biosolids recycle rate are summarized in Figures 12a-12e. For the control reactor (Reactor #1-Figure 12a), the actual anoxic HRT of 28 minutes was more than adequate for complete denitrification of all available nitrate+nitrite; 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 7 minutes of the anoxic phase was accompanied by orthophosphate release (Figure 12a).

The actual aerated 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 12a). Contrary to the results described earlier for the batch test investigation of denitrified recycle rate (Figure 11a), it appears that the pilot plant as it was operating on August 29 did not have enough reserve capacity to handle a daily peak flow of 1.65Q, at least over the short term.

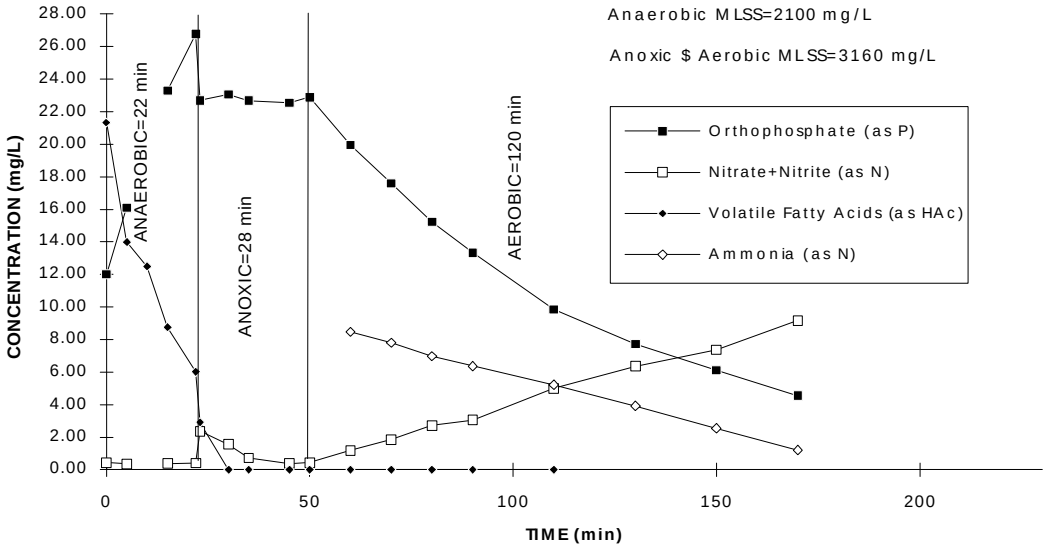


Figure 12a-Batch Reactor #1-Simulated Settled Biosolids Recycle=Q

The simulated effects of reducing the biosolids recycle from Q to 0.5Q (Reactor #2) are shown in Figure 12b. The decrease in recycle resulted in an increase in anoxic actual HRT from 28 minutes to 32 minutes, and an increase in aerobic HRT from 120 minutes to 147 minutes. However, the decrease in recycle also resulted in a decrease in anoxic and aerobic MLSS concentration to 2550 mg/L (Figure 12b), compared to 3160 mg/L in the control reactor (Figure 12a) . The net effect was that the concentration of orthophosphate in solution at the end of the aerobic phase in Reactor #2 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+nitrite 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.

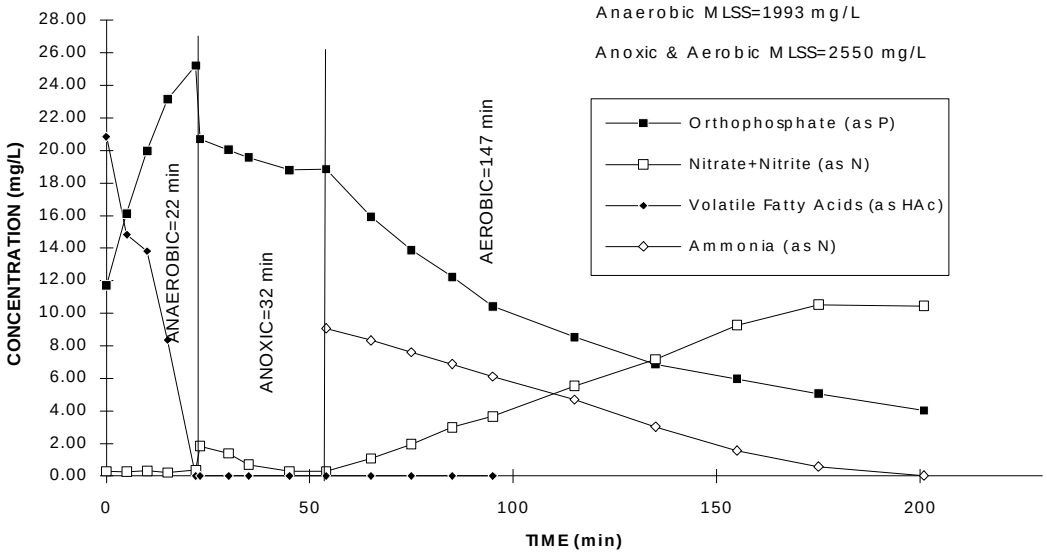


Figure 12b-Batch Reactor #2-Simulated Settled Biosolids Recycle=0.5Q

The simulated effects of reducing the biosolids recycle to 0.25Q (Reactor #3) are shown in Figure 12c. In this case, the simulated decrease in recycle increased the anoxic HRT to 34 minutes, and increased the aerobic HRT to 166 minutes. 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. Similar to Reactor #1, VFA were remaining in solution at the end of the anaerobic phase, and 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 increasing the biosolids recycle to 1.5Q (Reactor #4) are shown in Figure 12d. In this case, the anoxic and aerobic MLSS concentration was increased to 3660 mg/L (Figure 12d), compared to 3160 mg/L in the control reactor (Figure 12a), 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).

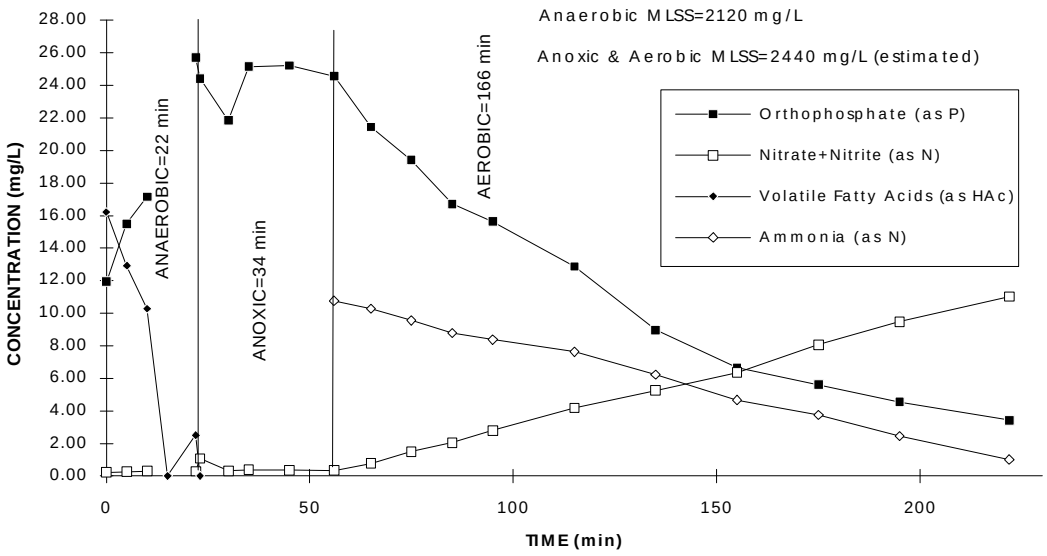


Figure 12c-Batch Reactor #3-Simulated Settled Biosolids Recycle=0.25Q

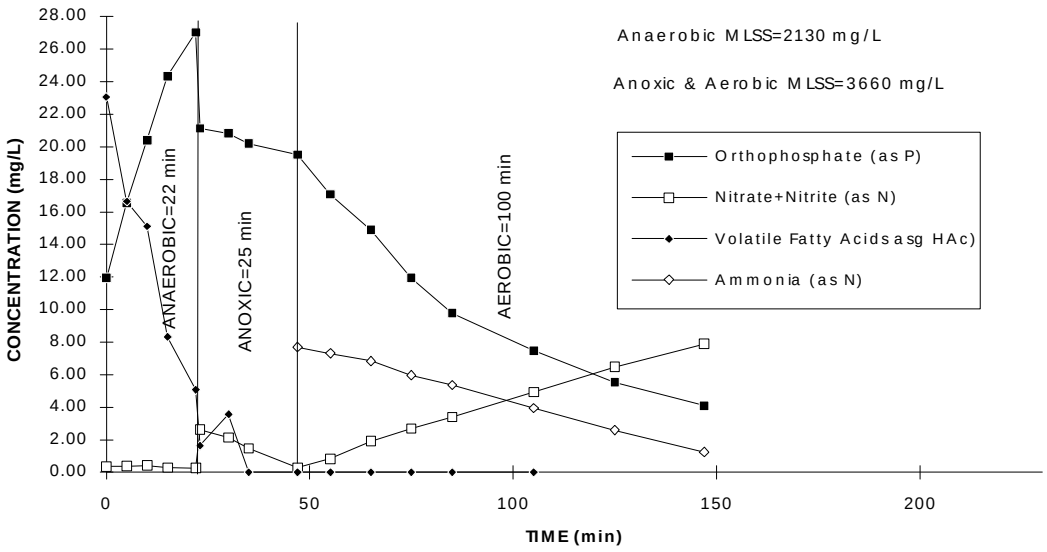


Figure 12d - Batch Reactor #4-Simulated Settled Biosolids Recycle=1.5Q

The effects of further increasing the biosolids recycle to 2Q (Reactor #5) are shown in Figure 12e. The anoxic and aerobic HRTs were reduced to 23 minutes and 87 minutes, respectively, and 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).

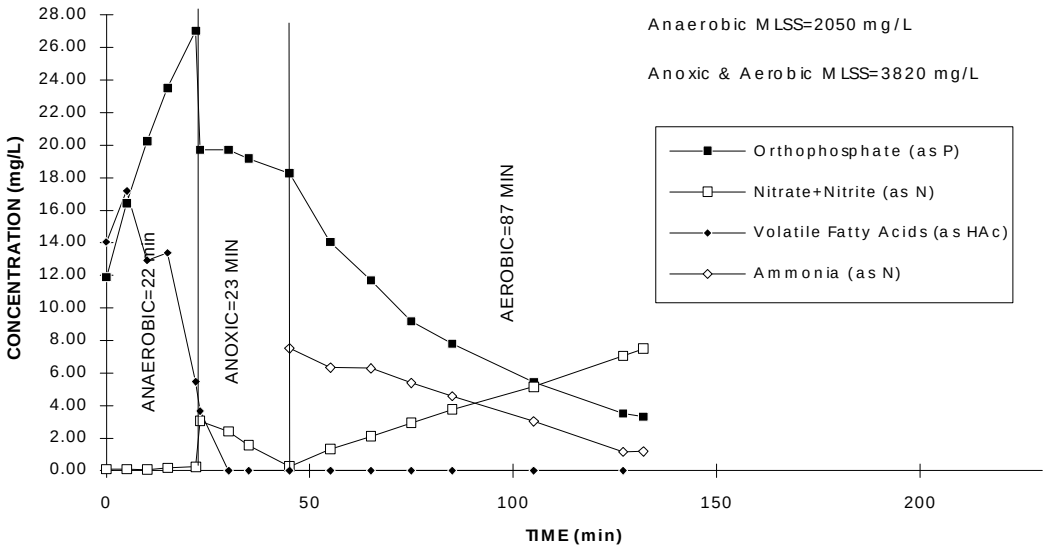


Figure 12e-Batch Reactor #5-Simulated Settled Biosolids Recycle=2Q

In summary, for the batch test simulation of manipulating the biosolids recycle, 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 (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). Decreasing the biosolids 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. It therefore appears 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.

7 TASK 3 SOLIDS RETENTION TIME

7.1 Pilot Plant Performance

The solids retention time (mean cell residence time) in an activated sludge reactor is defined as the mass of cells in the reactor (usually the aeration basin) divided by the mass of cells wasted from the reactor per day. The operating mixed liquor suspended solids (MLSS) concentration in the reactor is a function of the solids retention time (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; 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 solids retention time of the fixed growth in the FGR. 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 Tasks 1 and 2, 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. After the investigation into FGR recycle rates was completed, the sludge wasting rate on the B Side was increased in July of 1994, to reduce the operating aeration basin MLSS concentration as close to 2000 mg/L as possible, and the A Side was maintained at 3000 mg/L. All other parameters for both sides were kept at the levels described for the B Side Phase 3 in Table 3 earlier.

The investigation of optimum operating MLSS concentration (secondary sludge wasting rate) was carried out during the period from July 26 through August 31 of 1994. During this period, 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. The resulting average aeration basin operating MLSS concentration was 3090 mg/L on the A Side and 1970 mg/L on the B Side.

Process removals of total suspended solids, BOD_5, total kjeldahl nitrogen, and total phosphorus during the investigation into operating MLSS concentration are summarized in Table 8. The results of the *t* test (not shown in Table 8) indicated that the mean removal of total suspended solids was significantly higher on the B Side (79 mg/L) than on the A Side (63 mg/L), suggesting that the lower operating MLSS concentration on the B Side might have resulted in improved removal of suspended solids. However, the mean concentration of total suspended solids in the process influent on the B Side over the same period (90 mg/L) was significantly higher than on the A Side (75 mg/L). Mean final effluent concentrations of total suspended solids were not significantly different between the two sides (12 mg/L for the A Side and 11 mg/L for the B Side). Therefore, the significantly higher TSS removal observed on the B Side was likely due to the higher influent TSS, and not to the lower operating MLSS concentration.

Removals of total BOD_5 and total kjeldahl nitrogen were not significantly different between the two sides (Table 8). However, mean removal of total phosphorus was significantly higher on the A Side (3.6 mg P/L) than on the B Side (1.8 mg P/L).

Table 8 - Pilot Plant Performance During Task 3

Parameter	Mean Process Removal (mg/L)	
	A Side	B Side
Total Suspended Solids	63	79
Total BOD_5	93	97
Total Kjeldahl Nitrogen (as N)	21.3	19.4
Total Phosphorus (as P)	3.6	1.8

The results of mass balances conducted on the individual process reactors during Task 3 are summarized in Table 9. The mass balance removals were based on nominal retention times, rather than actual retention times. That is, actual removals were normalized to reflect the removal per litre of process influent flow, excluding the effects of recycle. The results of the *t* test (not included in Table 9) indicated that the increase in secondary sludge wasting rate (*i.e.*, the lower operating MLSS concentration) on the B Side did not significantly affect process ammonia removal or nitrate production. 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 (Table 9). However, the batch test results presented later in this Chapter showed that the suspended growth nitrification and denitrification rates were lower on the B Side than the A Side during this period.

Table 9 - Summary of Reactor Mass Balances During Task 3

Process Reactor	Mean Removed or Produced (mg/L of Process Influent)					
	Ammonia (as N)		Nitrate+Nitrite (as N)		Orthophosphate (as P)	
	A Side	B Side	A Side	B Side	A Side	B Side
Anaerobic Reactor	N/A[1]	N/A[1]	N/A[1]	N/A[1]	37.0 (released)	21.6 (released)
Anoxic Reactor	N/A[1]	N/A[1]	9.0 (removed)	7.1 (removed)	8.3 (removed)	7.4 (removed)
FGR Cell 1	7.8 (removed)	6.9 (removed)	11.2 (produced)	8.7 (produced)	14.8 (removed)	4.3 (removed)
FGR Cell 2	4.8 (removed)	4.6 (removed)	5.7 (produced)	6.0 (produced)	8.8 (removed)	4.7 (removed)
Aeration Reactor	N/A[1]	N/A[1]	N/A[1]	N/A[1]	7.1 (removed)	5.3 (removed)
Total Process	12.9 (removed)	12.7 (removed)	8.8 (produced)	8.4 (produced)	38.4 (removed)	20.9 (removed)

[1] Not Applicable

On the other hand, the difference in operating MLSS concentration between the two sides had a significant effect on process phosphorus removal. Mean release of orthophosphate in the A Side anaerobic reactor was 37.0 mg P/L, compared to only 21.6 mg P/L on the B Side. Corresponding mean anaerobic uptake of volatile fatty acids was 43.6 mg HAc/L on the A Side and 28.0 mg HAc/L on the B Side. According to the *t* test, both phosphate release and VFA uptake were significantly higher on the A Side. During the investigation of sludge wasting rate, mean VFA concentration in the process influent did not differ significantly between the A Side (mean=44.2 mg HAc/L) and the B Side (mean=41.6 mg HAc/L), suggesting that the observed difference in anaerobic VFA uptake was due to the controlled difference in operating MLSS concentration.

According to theory, lower anaerobic carbon (VFA) storage should lead to lower bacterial uptake and storage of phosphorus in the subsequent aerobic phase (see earlier discussion in Chapter 3). The data in Table 9 are in accordance with theory. In the anoxic reactor, denitrification with associated bacterial uptake of orthophosphate was not significantly different between the A and B Sides. However, 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). 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.

It is apparent from the data discussed above that the suspended biomass on the A Side had a much higher phosphate release and uptake capacity than that of the B Side, and that overall removal of phosphorus was significantly better on the A Side than on the B Side.

7.2 Batch Testing

Typical results of bench-scale batch testing to compare the performance of the A and B Sides during the investigation into optimum operating MLSS concentration are summarized in Figure 13. The batch test results confirm the results discussed above in Section 7.1. For the A Side (Figure 13a), 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 13b).

At the end of the aerobic phase, the final orthophosphate concentration on the A Side was 0.65 mg P/L, compared to 2.3 mg P/L on the B Side (Figure 13). It is apparent from the batch test results that the lower solids retention time achieved by the higher secondary sludge wasting rate (*i.e.*, the lower operating MLSS concentration) on the B Side resulted in much slower rates of bacterial phosphate release and uptake than on the A Side. The end result was a significant deterioration in biological phosphorus removal on the B Side, compared to the A Side.

Note that the rate of denitrification in the anoxic phase was higher on the A Side (Figure 13a) than on the B Side (Figure 13b); 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. The suspended nitrifiers in the FGR-SGR process were probably a mixture of fixed growth bacteria which sloughed off the FGR media to mix with the suspended growth, and bacteria which grew in suspension in the process mixed liquor.

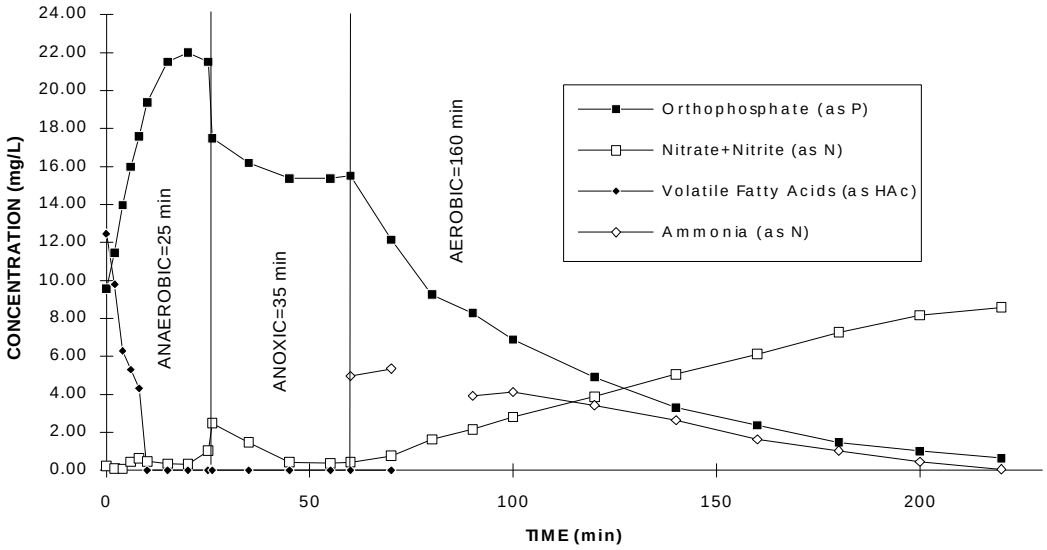


Figure 13a Task 3 Batch Test - A Side Process Train - August 25, 1994

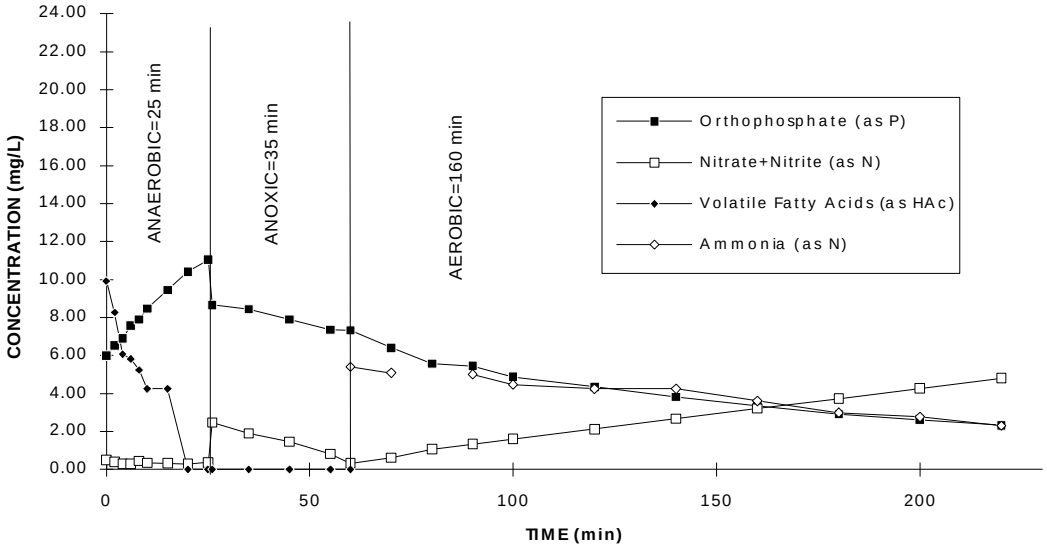


Figure 13b Task 3 Batch Test - B Side Process Train - August 25, 1994

In the FGR-SGR pilot plant, nitrification was mainly accomplished by the fixed growth bacteria in the FGRs, and very little ammonia reached the aeration

basin (see 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.

In light of the above, it is apparent that the lower sludge wasting rate which resulted in an operating MLSS concentration of 3090 mg/L resulted in the optimum system performance, compared to the higher sludge wasting rate, which resulted in an operating MLSS concentration of 1970 mg/L.

8 CONCLUSIONS

As described in Chapter 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.

- 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 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).

- 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 would reduce aeration requirements.

- Retention time in the anaerobic reactor should be sufficient to ensure bacterial uptake of all available volatile fatty acids (VFA) in the process influent. Allowing significant concentrations of VFA to reach the anoxic reactor tends to suppress phosphorus uptake during denitrification.

- The anoxic reactor should not be oversized (or, alternatively, under loaded with nitrate), or 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.

- Short-term daily increases in the FGR recycle rate to prevent solids accumulation on the media 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*).

- Manipulation of the denitrified recycle rate to dampen the effects of the peak daily hydraulic load typically experienced at full-scale plants had a detrimental effect on phosphorus removal over a single cycle. However, in light of current theory, it appears that increasing the denitrified recycle rate during the peak flow to attenuate the dilution of mixed liquor suspended solids in the anaerobic reactor might improve biological phosphorus removal over a large number of cycles, by increasing the degree of anaerobic carbon storage (based on bench-scale batch tests only).

- Manipulation of the settled biosolids recycle rate to dampen the effects of the peak daily load typically experienced at full-scale plants has the potential to improve phosphorus removal in the process. Increasing the simulated biosolids recycle flow rate by 50% to attenuate the dilution of mixed liquor suspended solids in the anoxic and aeration reactors had approximately the same beneficial effect on phosphorus removal as decreasing the recycle by 50% to increase the actual anoxic and aerobic hydraulic retention times (based on bench-scale batch tests only). However, decreasing the biosolids recycle would be the better course at a full-scale plant, since increasing the recycle might overload the final clarifier.

- Biological phosphorus removal, denitrification, and suspended-growth nitrification rates in the FGR-SGR process are higher at an operating aeration basin MLSS concentration of 3000 mg/L than at 2000 mg/L.

- 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 should allow the aeration basin to be sized for phosphorus removal only.

9 RECOMMENDATIONS FOR FURTHER RESEARCH

- Further work is required to investigate the long-term effects of manipulating the denitrified recycle flow rate and the return settled biosolids flow rate, in order to dampen the effects of the diurnal flow fluctuations typically experienced at full-scale treatment plants.
- The use of nitrates rather than dissolved oxygen for bacterial uptake and storage of phosphorus in biological nutrient removal systems should be further explored.
- The nitrification rate of the FGRs (trickling filters) in the FGR-SGR process should be investigated at higher ammonia loading rates.
- 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 effects of operating mixed liquor suspended solids concentration on biological phosphorus removal in the FGR-SGR process should be further investigated.
- The relative effects of volatile fatty acid concentration and hydraulic retention time in the anaerobic phase on biological phosphorus removal in the FGR-SGR process should be further investigated.
- The effects on phosphorus removal of allowing significant concentrations of volatile fatty acids to reach the anoxic reactor should be further investigated.
- 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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