Immersed Membrane Bioreactor Performance Evaluation:

Twelve Pines Sewage Treatment Plant

Suffolk County, New York

Final Report 04 -04
December 2004

New York State
Energy Research and
Development Authority

NYSERDA

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Final Report

Prepared for the

New York State
Energy Research and
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Albany, NY
www.nyserda.org

and

Twelve Pines Sewage Treatment Plant

Suffolk County, New York

Prepared by

O'Brien & Gere Engineers, Inc.

Syracuse, NY

Alan J. Saikkonen, P. E.
Damien R. Foster
Mark R. Greene, Ph. D.

and

Zenon Environmental Inc.

Oakville, Ontario
Washington, DC

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NOTICE

ABSTRACT

Increased public concern for health and the environment, the need to expand existing wastewater treatment plants due to population increases, and increasingly stringent discharge requirements, have created a need for innovative technologies that can generate high quality effluent at affordable cost. The membrane biological reactor (MBR) process is an innovative technology that warrants consideration as a treatment alternative where high quality effluent and/or footprint limitations are a prime consideration.

MBR processes have been applied for the treatment of industrial wastewaters for over ten years (Hare et al., 1990). In this process, ultrafiltration or microfiltration membranes separate the treated water from the mixed liquor, replacing the secondary clarifiers of the conventional activated sludge process. Historically, energy costs associated with pumping the treated water through the membranes have precluded widespread application for the treatment of high volumes of municipal wastewater. However, recent advancements in membrane technology, which have lead to reduced process energy costs, have induced wider application for municipal wastewater treatment (Thompson et al., 1998).

This report describes a pilot scale demonstration study conducted to test an MBR process for use in the Long Island Sound Drainage Basin.

- The pilot scale system demonstrated the ability of the process to achieve high levels of BOD_5 and ammonia removal efficiencies. The ability to achieve high levels of total nitrogen removal without the addition of a carbon source like methanol was also demonstrated for short periods of time. Many things including the complexity of the process, lack of a dedicated operator, equipment malfunctions, and the inability to operate within alarm conditions hampered sustained operation of the pilot system.

- An economic analysis of MBR processes vs. conventional processes (capable of achieving similar levels of total nitrogen removal) indicated that capital costs for a small MBR system (less than 0.5 MGD) may be approximately 10 – 15% more costly than a conventional system, and that annual operations and maintenance costs for a small system MBR system may be approximately 33% more expensive than a conventional system.

Key Words: Membranes, Membrane Bioreactor, Microfiltration, Nitrogen Removal, Ultrafiltration, Waste Water Treatment, ZENON

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SUMMARY

During the period from May 2001 through February 2002, a pilot test demonstration study was conducted to evaluate immersed membrane biological reactor (MBR) technology at the Twelve Pines Sewage Treatment Plant (STP) in Suffolk County, New York. The pilot study was conducted with primary effluent. The primary objective of the project was to verify that the MBR process was capable of achieving the necessary effluent quality goals. Total nitrogen removal (nitrification-denitrification) without supplemental carbon source addition was targeted in particular.

PROCESS DESCRIPTION AND OPERATIONAL ADVANTAGES

The MBR system is a technological advancement of the conventional biological treatment system (activated sludge) wherein the solids separation (clarification) process is replaced with ultrafiltration membranes. The hollow fiber membranes, which are immersed in the aeration tank (biological reactor), are connected to suction duty pumps, which apply a partial vacuum to the immersed hollow fibers to create a small pressure drop across the membrane surface. Clean, treated water passes through the membrane (0.04 micron pores) while biosolids are retained in the biological reactor. Excess biosolids are periodically wasted from the reactor, such that a relatively stable quantity of biomass is maintained in the reactor.

The MBR process produces a high quality, treated effluent equivalent to the combination of conventional activated sludge treatment followed by sand filtration. The MBR process will generally require a significantly smaller biological reactor tank than conventional treatment systems. The MBR process is less vulnerable to process upsets and biomass washouts during high wet weather flows. Additionally, the MBR process is better able to economically achieve ammonia and nitrogen removal in cold weather, as the MBR system has the ability to operate with a higher biomass concentration than conventional systems.

MEMBRANE PERFORMANCE

The membrane performance throughout the study was exceptional. The data collected shows no breach of membrane integrity, as 96% of the measurements had turbidity values less than 0.1 NTU.

During the majority of the study, the pressure difference across the membrane in the MBR system was less than 4 psi. Maintenance cleaning done by aerating the membranes was conducted weekly for the first few months of the study and as required during the final months of the study. On a number of occasions, the system shut down due to the high vacuum alarm, which would be triggered when the pressure differential across the membrane climbed due to the deposition of biosolids on the membrane surface (i.e., fouling due to solids accumulation). In each instance, aerating the membrane for several hours and conducting a maintenance cleaning decreased the required vacuum to an acceptable level.

One "recovery cleaning" was conducted at the end of the study. Cleaning the membranes with chlorine at 200 mg/L did little to improve the permeation rate; however soaking the membrane in 1,000 mg/L of citric acid restored the membrane permeability to its original state.

PILOT SYSTEM PERFORMANCE

The MBR pilot system did a very good job of removing all the BOD_5 and ammonia from the influent wastewater, which was supplied from the primary effluent stream at the STP. The pilot system had difficulty achieving the total nitrogen removal goal without the addition of methanol to assist in the denitrification process. The goal was achieved for short periods, but sustained operation with satisfactory total nitrogen removal performance was not achieved. Only one phase of the test program was completed, the one involving operation to measure the lowest total nitrogen removal without using methanol (or another carbon source) to facilitate denitrification. The additional planned phases were not completed due to the length of time it took to get reliable operation to complete the first phase of the program. However, information from other pilot and full scale MBR systems was gathered to show the performance of this technology under the operating conditions planned for the subsequent phases of the test program.

During the study, permeate quality was affected by a number of system shut downs and process upsets. However, when the system was operating within the targeted operational parameters, the effluent quality was very good, with permeate ammonia-nitrogen less than 1 mg/L and BOD_5 less than 5 mg/L.

A mixed liquor suspended solids (MLSS) concentration of 8,000 to 10,000 mg/L in the Membrane Tank was targeted, however, the actual MLSS readings fluctuated between 1,000 and 27,000 mg/L.

The ability of the MBR to achieve high levels of total nitrogen removal without the addition of a carbon source like methanol was also demonstrated for short periods of time. Many things including the complexity of the process, lack of a dedicated operator, equipment malfunctions, and the inability to operate within alarm conditions hampered sustained operation of the pilot system. Operating data acquired from other full scale MBR systems does demonstrate that high levels of TN removal may be achieved with this technology when using methanol as a carbon source for denitrification.

ECONOMIC EVALUATION

An economic analysis comparing the MBR process with a conventional process that used effluent filtration (i.e., systems capable of achieving similar levels of total nitrogen removal with carbon addition) was prepared. The analysis indicated that capital costs for a small MBR system (less than 0.5 MGD) may be approximately 10 to 15% more costly than a conventional system, and that annual operations and maintenance costs for a small MBR system may be approximately 33% more expensive than a conventional system. Since it appeared that methanol addition would be necessary to achieve the targeted total nitrogen removal performance, the economic analysis assumed this consumption would be similar for both treatment systems and therefore, costs associated with methanol addition were not included in the analysis.

In a typical municipal wastewater treatment facility, the biological treatment process (MBR or conventional) normally represents approximately 25% of the total plant's capital cost and approximately 30 to 40% of the plant's annual operations & maintenance costs.

CONCLUSIONS

The following conclusions can be drawn from this pilot study:

- MBR system permeate (effluent) ammonia-nitrogen levels of less than 1 mg/L were easily achieved when appropriate operating parameters were maintained.
- MBR system permeate (effluent) BOD_5 levels were consistently less than the study goal of 5 mg/L, when the system was operating within appropriate parameter ranges and healthy microorganisms were maintained.
- A recovery cleaning did not have to be conducted on the membranes until the system had operated for nine months. Regular maintenance cleaning and proper aeration of the membranes resulted in a recovery cleaning interval greater than the normal, manufacturer recommended period of six months.
- Total nitrogen levels of less than 8 mg/L in the permeate were achievable for short periods, albeit not consistently, without chemical addition.
- Total nitrogen levels of less than 8 mg/L have been successfully achieved at other full scale MBR operating installations with the use of methanol for denitrification.
- An economic analysis indicates that MBR systems can cost approximately 10 to 15% more to construct and approximately 33% more to operate than a conventional biological treatment systems using effluent filtration.
- The inability of the pilot unit to attain proper and reliable process operating conditions during portions of this study due to a variety of reasons needs to be addressed before conducting further studies with this particular equipment.

Section 1
INTRODUCTION

The New York State Energy Research and Development Authority (NYSERDA), together with O'Brien and Gere Engineers, Suffolk County and ZENON Environmental Systems Inc. (Zenon), conducted a pilot test study to evaluate immersed membrane biological reactor technology at the Twelve Pines Sewage Treatment Plant (STP) in Suffolk County, New York.

The purpose of the membrane biological reactor (MBR) pilot plant study was to assess the ability of the process to produce a high quality effluent, targeting nitrogen removal in particular. Total nitrogen (TN) removal is of importance to the Twelve Pines STP and other STPs in Suffolk County because these plants discharge to aquifers via recharge basins.

In April 2001, a pilot scale immersed ultrafiltration membrane bioreactor was delivered to the site by Zenon. The study was conducted over an eleven month period commencing in May 2001 and operating until March 2002.

Section 2
OBJECTIVES

The main goal of the pilot program was to demonstrate performance of the MBR process in the treatment of municipal wastewater, especially in the removal of total nitrogen without adding a carbon source like methanol.

The pilot objectives included:

- demonstrating that the MBR process could reliably and consistently produce a permeate (effluent) meeting or surpassing current effluent discharge standards.
- determining the lowest achievable total nitrogen level in the permeate without methanol addition.
- determining the lowest achievable total nitrogen level in the permeate with methanol addition.
- determining the lowest methanol dose required to achieve and maintain total nitrogen levels or less than 8 mg/L.
- demonstrating nitrogen removal with cold temperature feed water.
- conducting a membrane integrity test upon completion of the pilot activities.
- Meeting the following permeate (effluent) concentration limits:
  - CBOD5 <5 mg/L
  - TSS <1 mg/L
  - TN (total nitrogen) <8 mg/L

Subsequent to the completion of pilot operations and evaluation of operating data, an economic evaluation was prepared. The economic evaluation compares the capital and operating costs of an MBR system to that of a conventional system with effluent filtration.

Section 3
PROCESS DESCRIPTION

SYNOPSIS OF THE MEMBRANE BIOREACTOR WASTEWATER TREATMENT PROCESS

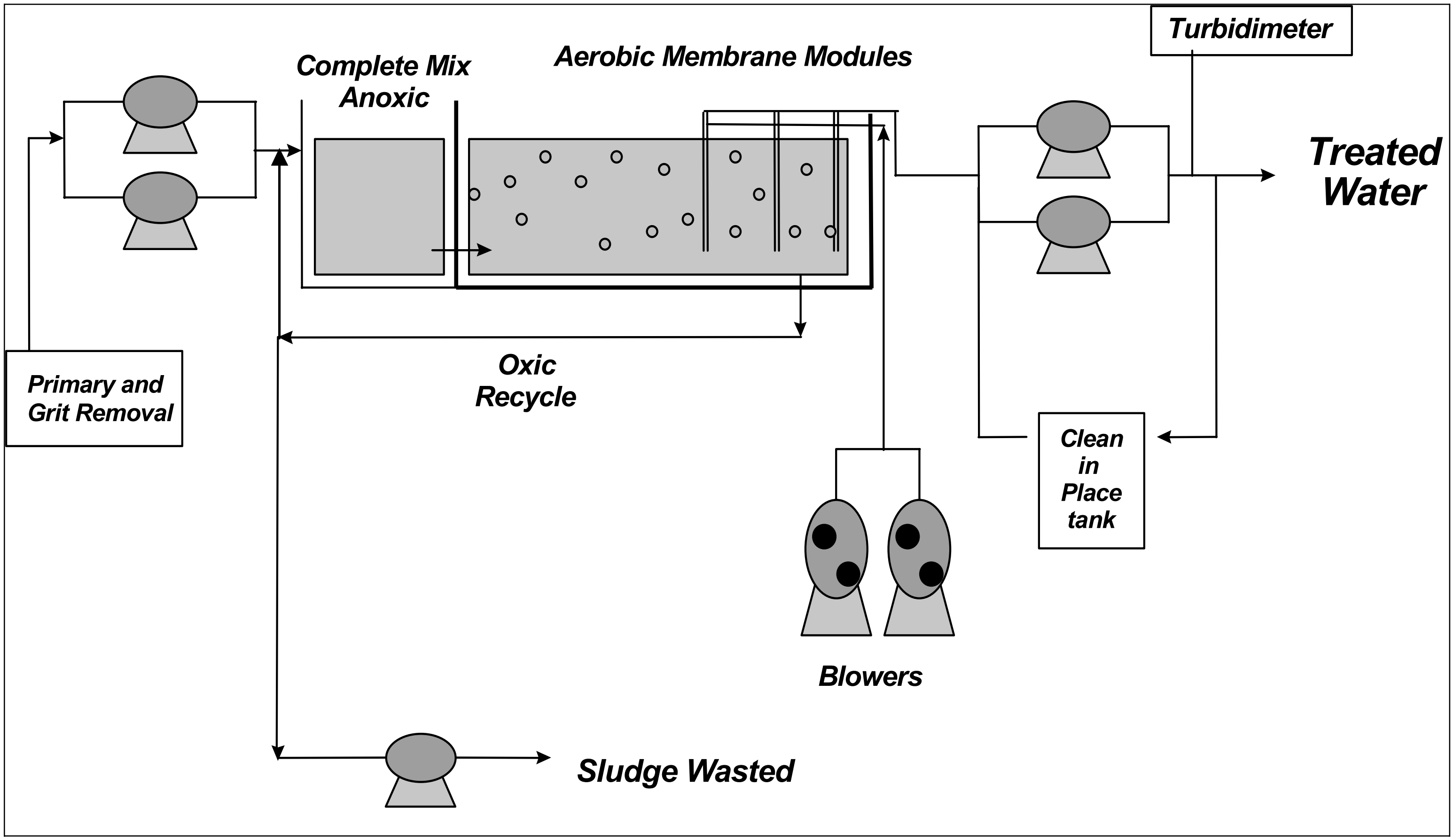
The MBR process technology consists of a suspended growth biological reactor integrated with an ultrafiltration membrane system. Figure 3-1 is a process flow schematic of the MBR process used for carbonaceous removal and nitrification/denitrification. Essentially, the ultrafiltration system replaces the solids separation function of a conventional activated sludge system (secondary clarifiers and sand filters). For municipal wastewater applications, the membrane filter consists of hollow fiber material with a 0.04 micron nominal pore size. This pore size precludes the passage of particulate material from being discharged with the effluent.

The membranes are typically submerged in the aeration tank, in direct contact with the mixed liquor. Through the use of a suction duty pump, a vacuum is applied to a header connecting the membranes. The vacuum draws the treated water through the membranes. The use of a vacuum, rather than positive pressure, greatly reduces the energy associated with permeate pumping. Air is intermittently introduced to the bottom of the membrane modules through integrated coarse-bubble diffusers. This produces turbulence which scours the external surface of the hollow fibers transferring rejected solids away from the membrane surface. This aeration also provides the required oxygen necessary for the biological process to flourish. Waste sludge is periodically pumped from the aeration tank, such that a relatively constant MLSS concentration is maintained.

The MBR process effectively overcomes the problems associated with poor settling of biomass and loss of biomass to the effluent that can plague conventional activated sludge processes with gravity clarification. The MBR process permits bioreactor operation with considerably higher mixed liquor solids concentration than conventional activated sludge systems, which are limited by biomass settleability. The MBR process is typically operated at a MLSS concentration in the range of 8,000 to 12,000 mg/L whereas conventional activated sludge processes operate at approximately 1,000 to 3,000 mg/L MLSS. The elevated biomass concentration allows for highly effective removal of both soluble and particulate biodegradable material in the waste stream. The MBR process combines the unit operations of aeration, secondary clarification, and filtration into a single process, simplifying operation and greatly reducing space requirements.

Since the MBR process can be operated at elevated MLSS concentrations, extended solids retention times (SRT) are readily attainable. Accurate SRT control is very simple since no solids are lost via the effluent. Many municipal MBR plants are operated with a SRT exceeding 20 days. These extended SRTs ensure complete nitrification even under cold weather operating conditions. At extended SRTs, sludge yields can

Figure 3-1 Typical MBR Process Flow Schematic



be considerably less than conventional activated sludge process processes, due to endogenous decay of the biomass.

MBR IMMERSED MEMBRANE BIOREACTOR PILOT SYSTEM EQUIPMENT DESCRIPTION

The immersed membrane bioreactor system supplied to the Twelve Pines STP consisted of a permeate pump, membrane tank, blower, permeate recycle mixed liquor re-circulation equipment, anoxic and aerobic tanks. The system was supplied by ZENON Membrane Products, along with the necessary instrumentation and controls required for operation. The major components are summarized in Table 3-1.

Table 3-1 Twelve Pines STP MBR Pilot System Summary

Membrane manufacturer and place of manufacture	ZENON Environmental Inc, Burlington, Ontario
Size of membrane element used in study	6.8 ft x 2.5 ft x 0.7 ft (HxLxW)
Active membrane area of cassette used in study	660 ft^2
Membrane Pore size	0.04 µm (nominal)
Membrane material / construction	Proprietary Polymer
Membrane hydrophobicity	Hydrophilic
Membrane charge	Neutral
Design flux at the design pressure (GFD)	5 to 30 GFD
Acceptable range of operating pressures	-1 to -10 psi
Range of operating pH values	5 – 9.5
Range of Cleaning pH	2 – 11 (<30°C); 2 – 9 (>30°C)
Maximum concentration for OCl^- cleaning	2,000 ppm

Figure 3-2 shows a diagram of the pilot plant layout for the period of April 10 to August 26, 2001. Samples were collected from locations 1, 2 and 3 for determination of the performance of the system during the demonstration. Figure 3-3 is a process flow schematic for the pilot layout shown in Figure 3-2. There were two sets of aerobic and anoxic zones and two recirculation loops, one for each aerobic-anoxic pair of zones.

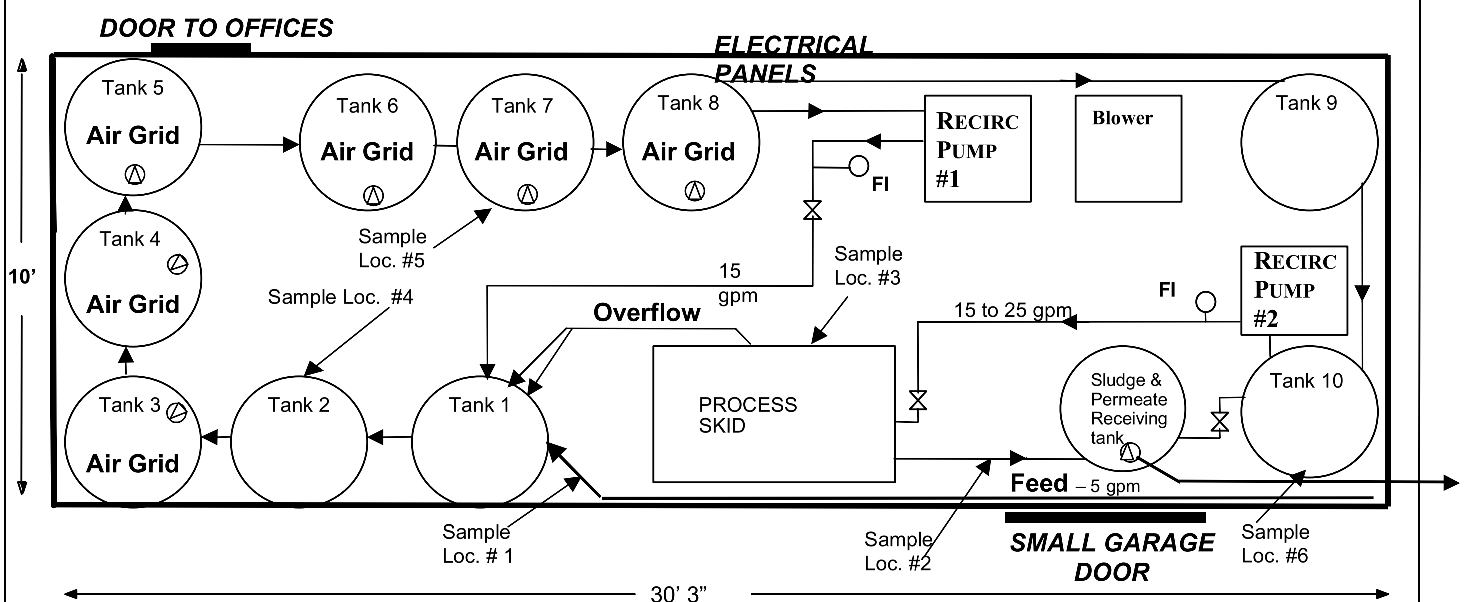
The configuration of the anoxic and aerobic tanks were changed twice during the study. Figure 3-4 is the pilot layout after the first change and this configuration was used from August 26 to November 7, 2001. Basically, Tank 8 was converted to anoxic operation and the overflow from the Membrane Tank was re-routed to Tank 3. Figure 3-5 is the process flow schematic for the layout shown in Figure 3-4. Later, it was found that the overflow from the Membrane Tank had two outfall connections and the second configuration change was to rectify this situation by re-routing the second connection to Tank 3.

The second configuration change is shown in Figures 3-6 (layout) and 3-7 (process schematic). In this configuration, the influent wastewater was passed through an anoxic zone before it was combined with the overflow from the Membrane Tank and sent to the aerobic zone.

Figure 3-2 Twelve Pines STP MBR Demonstration Layout (April 10 – August 26, 2001)

Sample Location Descriptions:

1. Influent (Feed Line)
2. Effluent (Permeate Line)
3. Membrane Tank (Process Skid/Aerobic Zone #2)
4. Last Stage of Anoxic Zone #1 (Tank #2 sample port)
5. Last Stage of Aerobic Zone #1 (Tank #7 sample port)
6. Last Stage of Anoxic Zone #2 (Tank #10 sample port)



*Note: In general only influent, effluent and membrane tank parameters will be measured (locations #1, 2 &3). During process optimization samples may be taken from locations 4 through 6.

1. FEED PUMP: 150 ft. away and down 8 ft. with an in-line basket strainer, pumped from center of primary clarifier.
2. WASTE SLUDGE: gravity feed to sludge holding tank then pumped to primary clarifier influent channel.
3. PERMEATE: discharged to sludge holding tank then pumped to primary clarifier influent channel.
4. CLEAN WATER SUPPLY: 60 psig tap water.

Figure 3-3 Twelve Pines STP MBR Process Flow Schematic (April 10 – August 26, 2001)

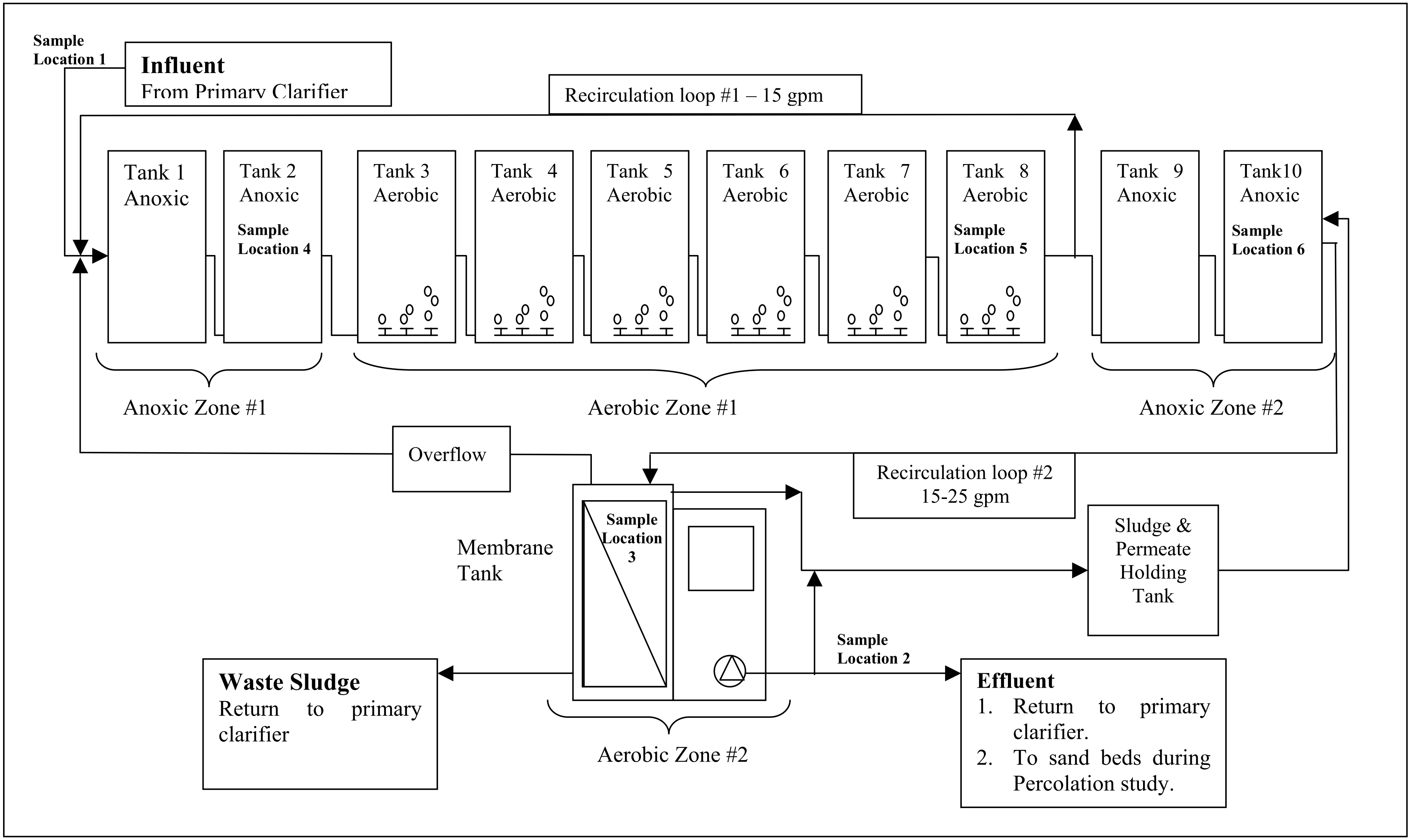
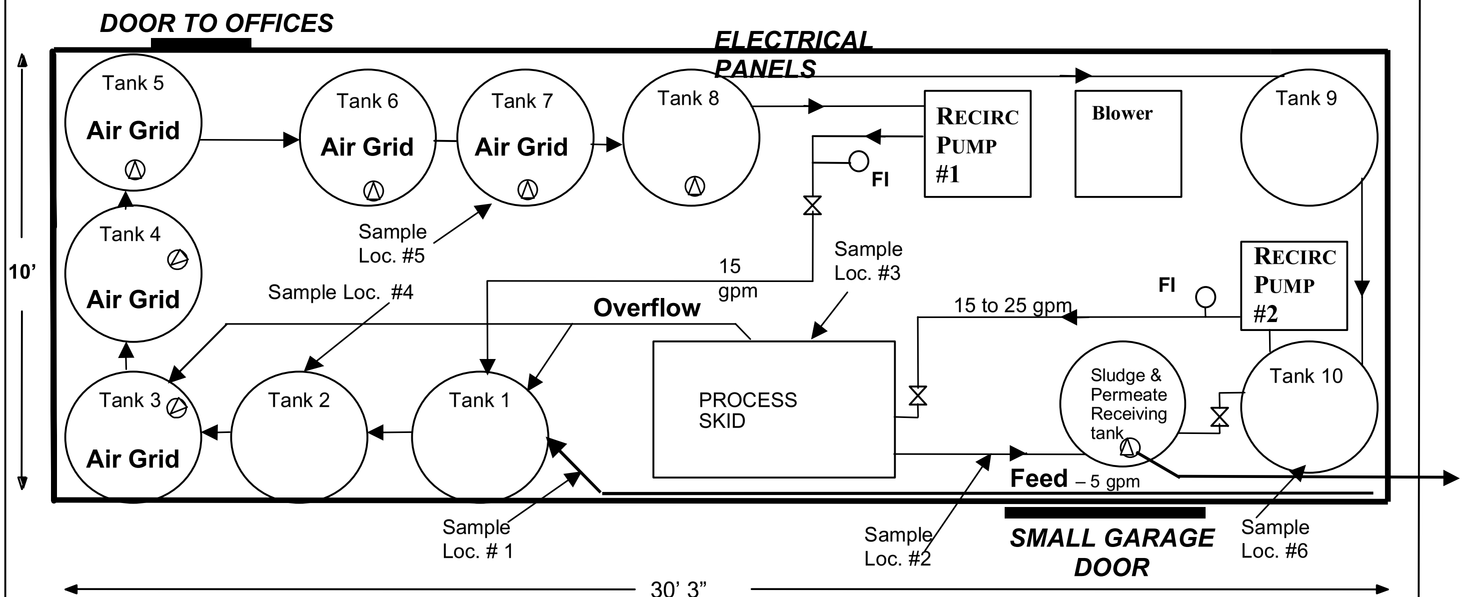


Figure 3-4 Twelve Pines STP MBR Demonstration Layout (August 26 – November 7, 2001)

Sample Location Descriptions:

1. Influent (Feed Line)
2. Effluent (Permeate Line)
3. Membrane Tank (Process Skid/Aerobic Zone #2)
4. Last Stage of Anoxic Zone #1 (Tank #2 sample port)
5. Last Stage of Aerobic Zone #1 (Tank #7 sample port)
6. Last Stage of Anoxic Zone #2 (Tank #10 sample port)



*Note: In general only influent, effluent and membrane tank parameters will be measured (locations #1, 2 &3). During process optimization samples may be taken from locations 4 through 6.

1. FEED PUMP: 150 ft. away and down 8 ft. with an in-line basket strainer, pumped from center of primary clarifier.
2. WASTE SLUDGE: gravity feed to sludge holding tank then pumped to primary clarifier influent channel.
3. PERMEATE: discharged to sludge holding tank then pumped to primary clarifier influent channel.
4. CLEAN WATER SUPPLY: 60 psig tap water.

Figure 3-5 Twelve Pines STP MBR Process Flow Schematic (August 26 – November 7, 2001)

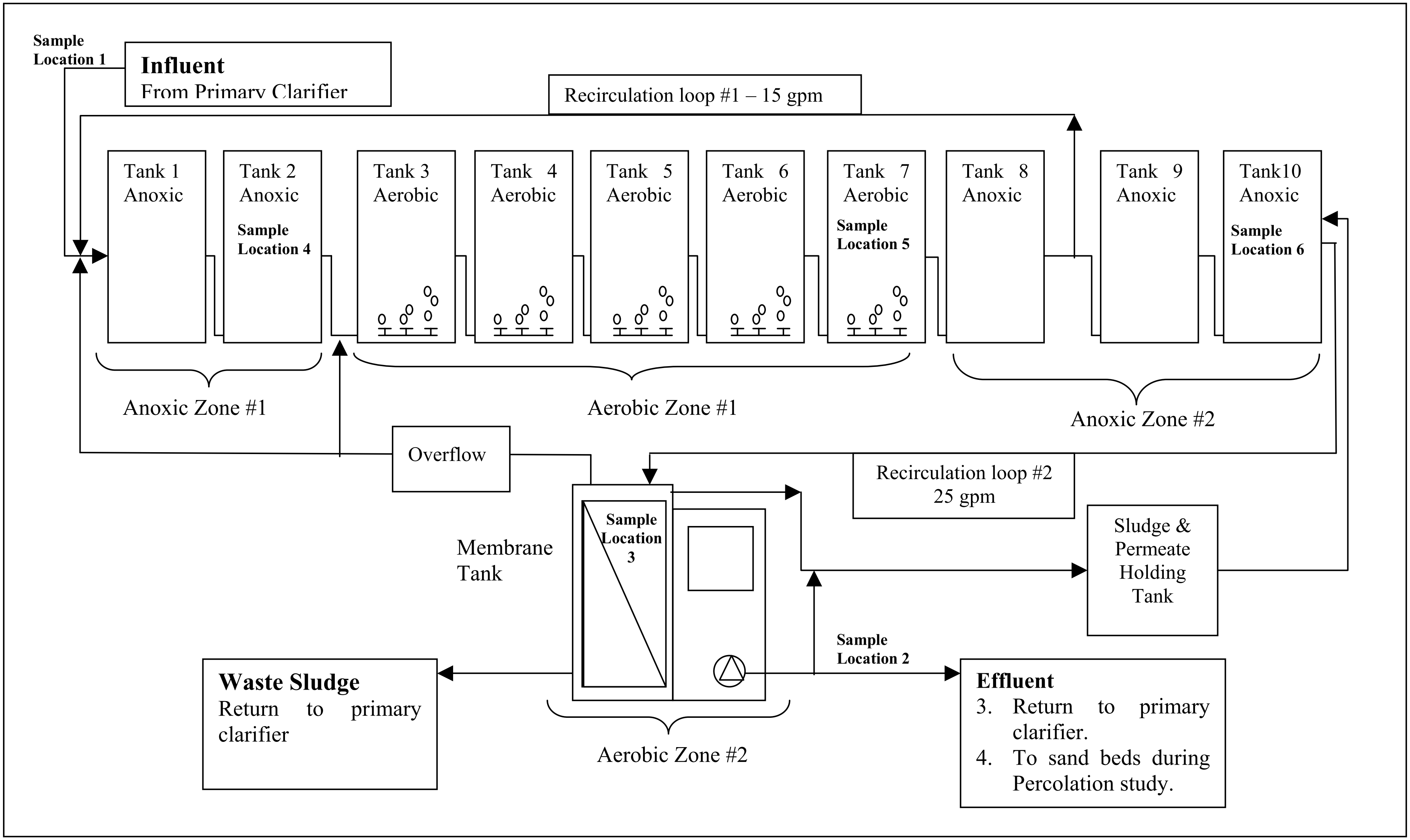
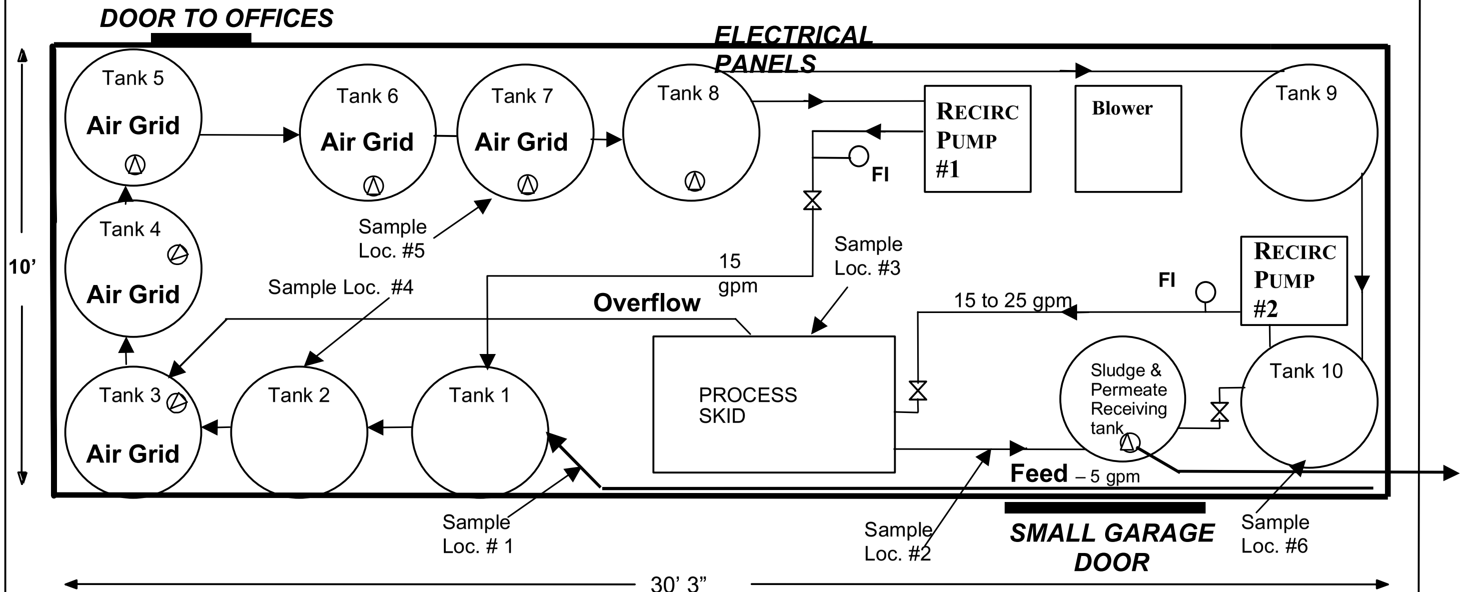


Figure 3-6 Twelve Pines STP MBR Demonstration Layout (November 7, 2001 – February 7, 2002)

Sample Location Descriptions:

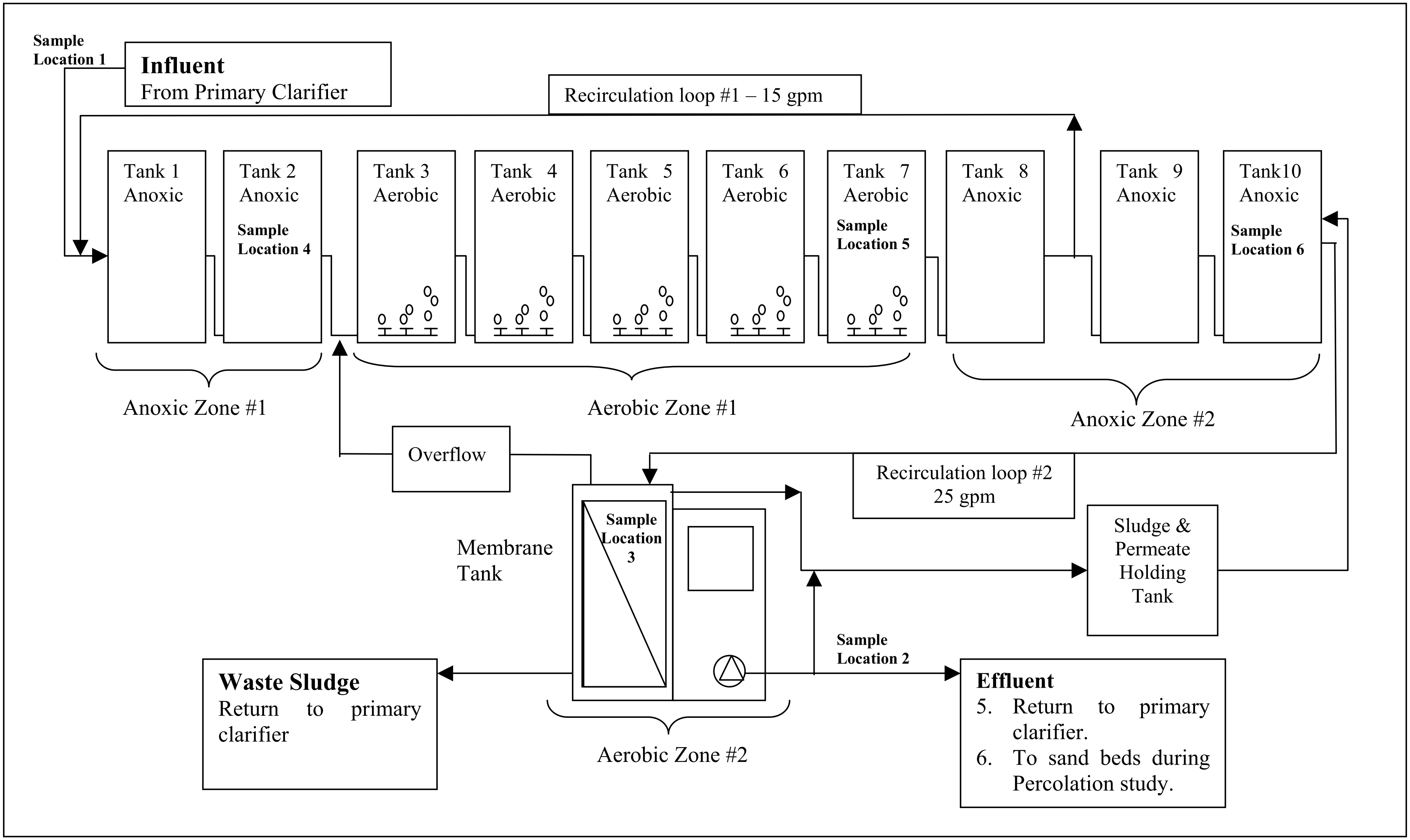
1. Influent (Feed Line)
2. Effluent (Permeate Line)
3. Membrane Tank (Process Skid/Aerobic Zone #2)
4. Last Stage of Anoxic Zone #1 (Tank #2 sample port)
5. Last Stage of Aerobic Zone #1 (Tank #7 sample port)
6. Last Stage of Anoxic Zone #2 (Tank #10 sample port)



*Note: In general only influent, effluent and membrane tank parameters will be measured (locations #1, 2 &3). During process optimization samples may be taken from locations 4 through 6.

1. FEED PUMP: 150 ft. away and down 8 ft. with an in-line basket strainer, pumped from center of primary clarifier.
2. WASTE SLUDGE: gravity feed to sludge holding tank then pumped to primary clarifier influent channel.
3. PERMEATE: discharged to sludge holding tank then pumped to primary clarifier influent channel.
4. CLEAN WATER SUPPLY: 60 psig tap water.

Figure 3-7 Twelve Pines STP MBR Process Flow Schematic (August 26 – November 7, 2001)



Major components of the MBR pilot include the following:

- Bag Filter Housing with 2mm screen
- Anoxic Tanks (4 through August 27 and 5 after August 27) (each tank volume 317 gallons)
- Aerobic Tanks (6 through August 27 and 5 after August 27) (each tank volume 317 gallons)
- Membrane Tank (total tank volume 185 gallons)
- Membrane & Supplemental Aeration Blowers
- Permeate Pump
- Sludge and Permeate Receiving Tank (total tank volume 100 gallons)
- One (1) MBR pilot membrane cassette
- CIP tank (25 gallons)
- Air compressor
- 2 horsepower submersible feed grinder pump
- Instrumentation and Controls
- Portable D.O. Meter
- On-line permeate turbidimeter

Section 4
OPERATIONAL AND ANALYTICAL PARAMETERS

OPERATIONAL PARAMETERS

The operational parameters for evaluating the performance of the MBR system are:

- permeate flux;
- vacuum pressure,
- permeability relaxation/backpulsing, and
- air scouring.

These parameters are described below.

Flux

Flux (also referred to as instantaneous flux) is a measure of the rate at which the product (or permeate) passes through the membrane per unit of surface area for the outside membrane surface. For an MBR process designed to provide biological wastewater treatment, permeate would be the system effluent. Flux is reported in units of liters per square meter per hour (LMH) or U.S. gallons per square foot per day (GFD). Net flux takes into account the production time lost during relaxation/backpulsing, and maintenance cleaning. Net flux also accounts for the actual volume of permeate lost during backpulsing. Instantaneous flux does not account for down time, and will always be a higher value than net flux.

Figure 4-1 shows the instantaneous flux and the net permeate flux during the operation of the demonstration. The instantaneous flux throughout the pilot study was maintained at 11 GFD.

Vacuum

Vacuum refers to the transmembrane pressure required to pull clean water through the membrane. Vacuum is reported in units of pounds per square inch (psi). The MBR system is designed to maintain a constant flux. Therefore, as the membrane becomes fouled, the transmembrane pressure increases. A cleaning is typically required once the transmembrane pressure exceeds 8 psi (vacuum) for an extended period of time.

Figure 4-2 shows the transmembrane pressure difference in psi. The vacuum pressures before and after backpulsing operations are plotted. As discussed below, backpulsing is a means of reducing the pressure drop across the membrane, and Figure 4-2 corroborates this effect. Over the course of the study, the system vacuum was not consistently recorded, however high vacuum alarms were noted on several occasions. Aerating the membranes restored the system vacuum on each occasion.

Figure 4-1: Instantaneous and net flux.

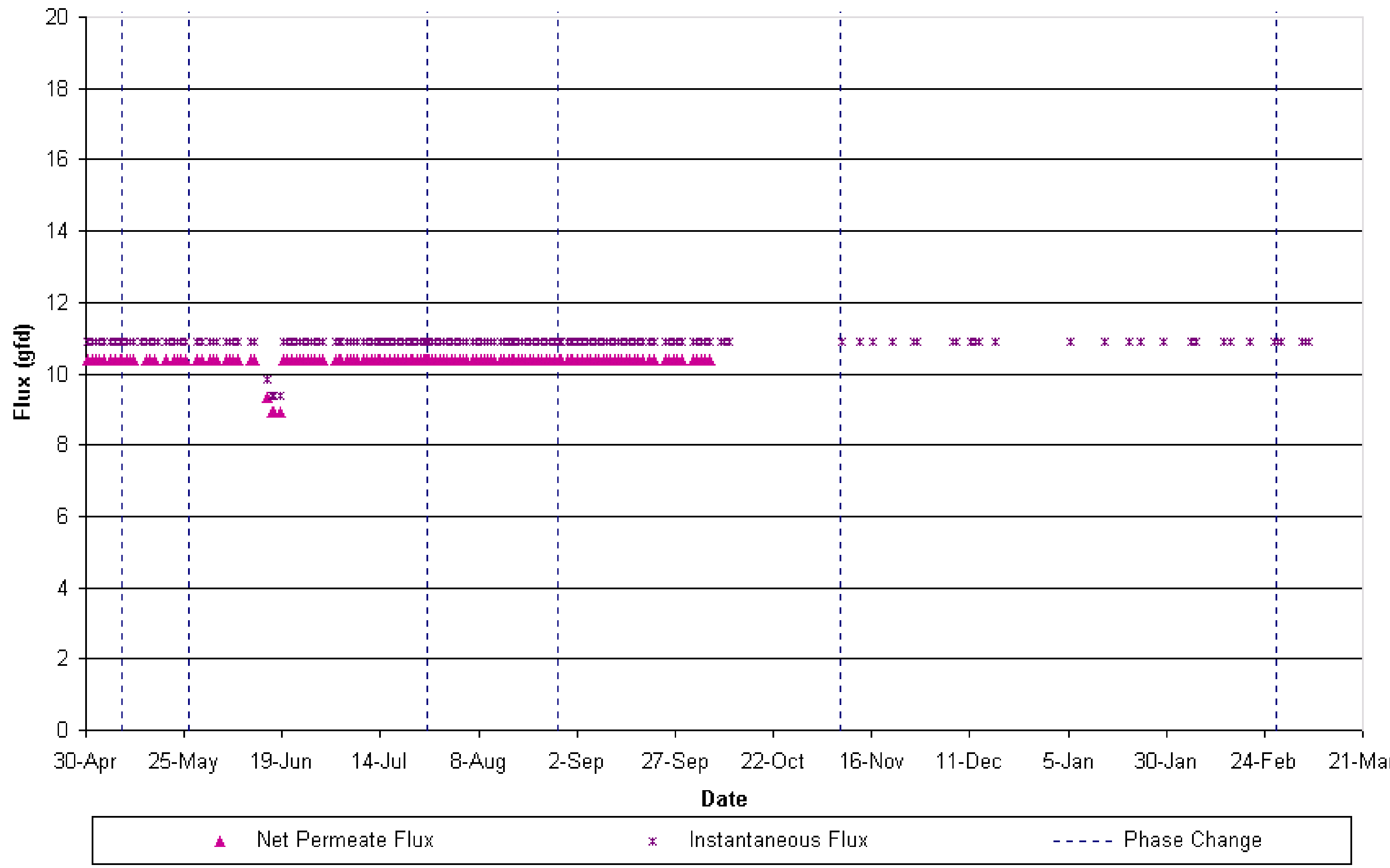
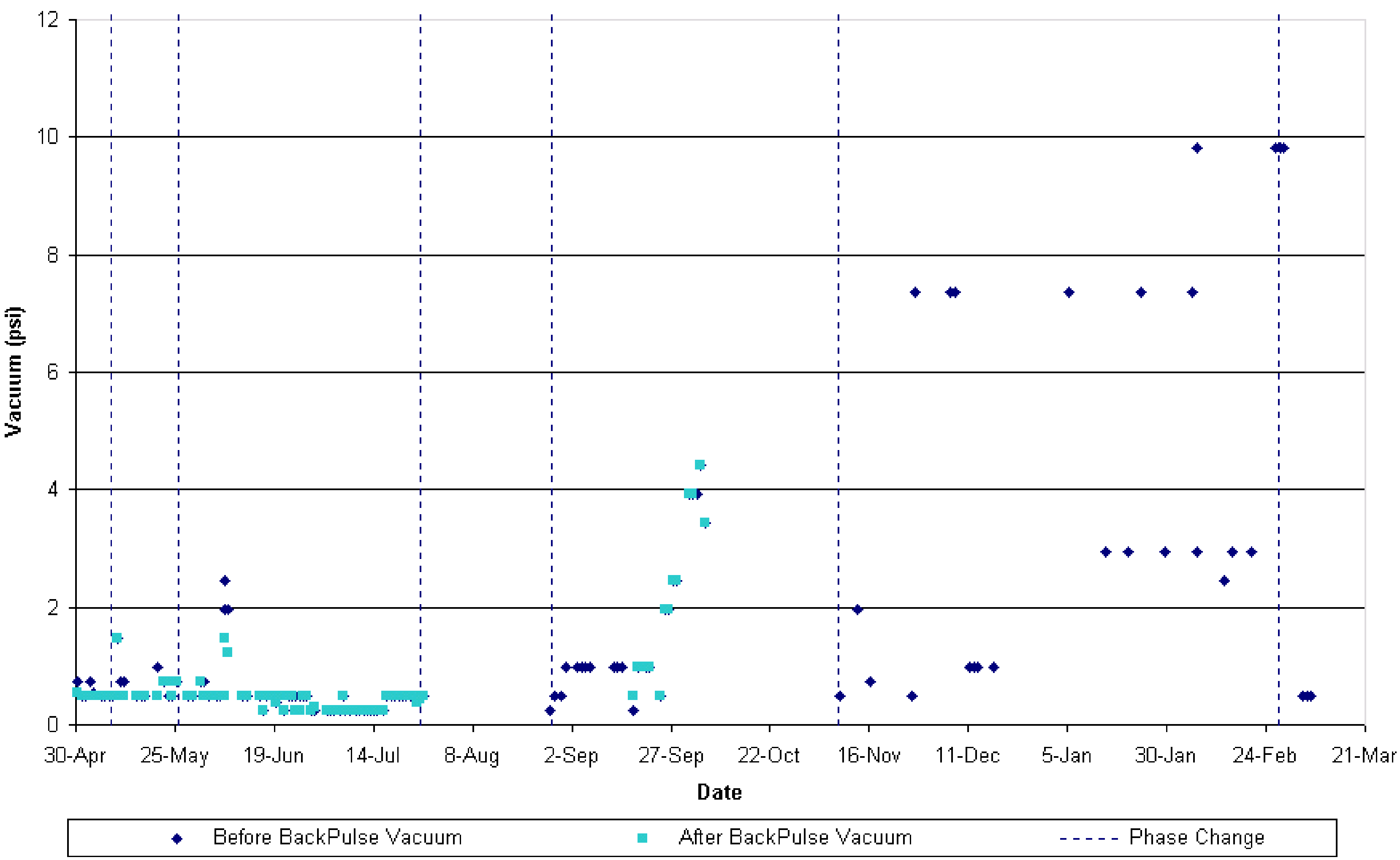


Figure 4-2: Before and after backpulse vacuum.



During the majority of the study, maintenance cleaning was conducted twice per week with sodium hypochlorite. One recovery clean was conducted at the end of the study (reference the cleaning discussion, for more detail).

Permeability

Permeability is a calculated parameter of flux normalized by transmembrane pressure. It is reported in units of GFD/psi. Permeability is typically corrected to account for temperature variations. Adjusting the permeability for temperature allows the influence of fouling to be determined. The formula used to calculate permeability at 20°C is based on the variance of the viscosity of water with temperature.

$$\text{Permeability @ } 20^{o}\text{C} = \text{Permeability @T x } 1.025^{(20-T)}$$

Figure 4-3 displays permeability and temperature. The permeability ranged from 1.1 to 22.2 GFD/psi for most of the study, while the temperature ranged from 16 to 27°C.

Relaxation and Backpulsing

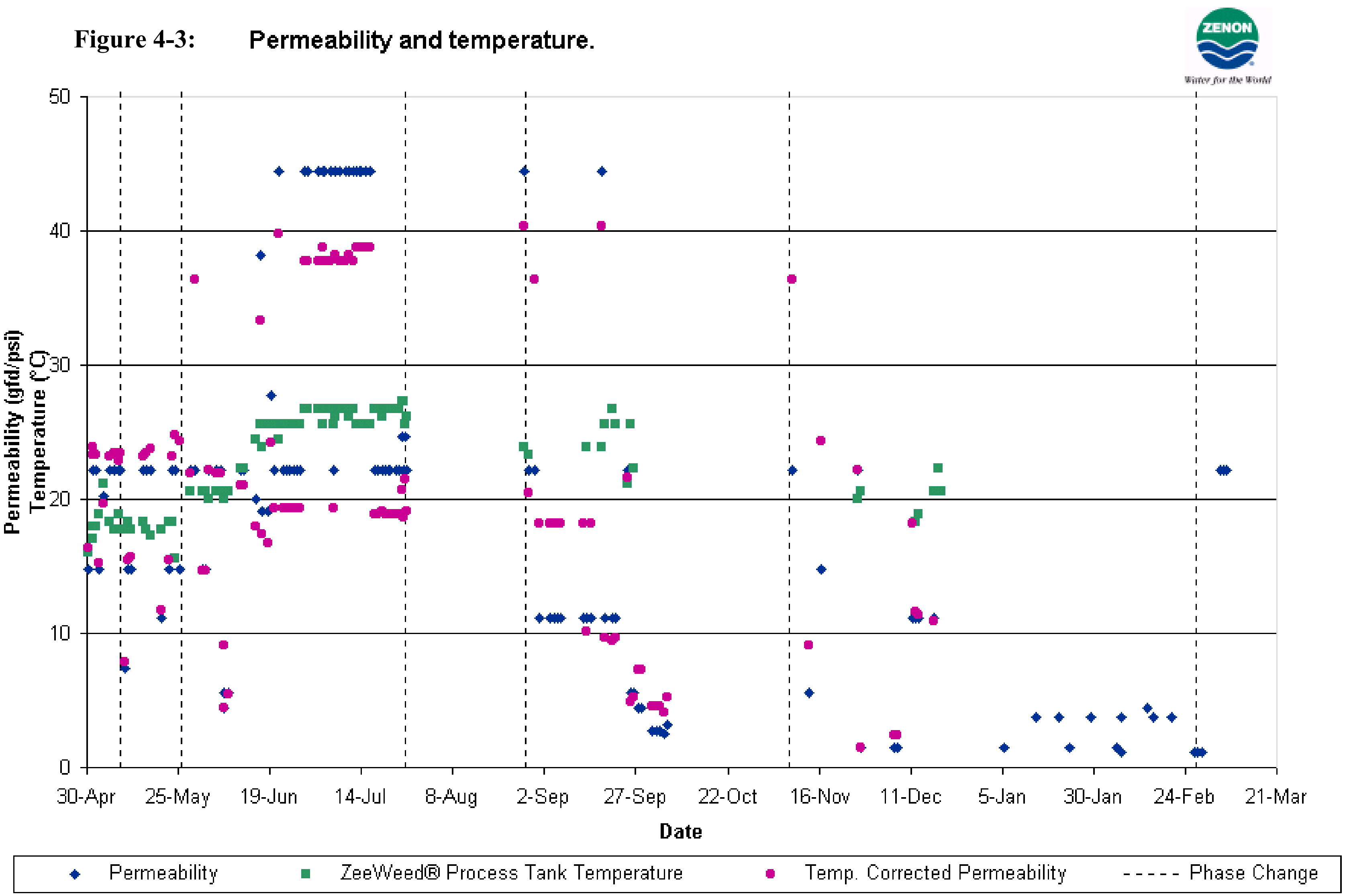
Relaxation is one component of the cleaning process. Every 10-20 minutes, flow through the membrane is stopped for 10-30 seconds. Relaxation frequency and duration should be optimized to extend the time between cleaning intervals or to increase production.

Air scouring is used to dislodge the cake layer on the membrane surface and to de-concentrate the solids within the membrane bundle during the relaxation period. In-house and field tests conducted by ZENON suggest that the major resistance to filtration in mixed liquor is the result of solids accumulation on the membrane surface. Air scouring in conjunction with relaxation has proven to be as effective as air scouring with backpulse (reversing the flow of permeate through the membranes). By replacing backpulse with relaxation, significant savings can be generated. Specific advantages of relaxation vs. backpulse include:

- Increased productivity - Net production with relaxation is 5-8% higher than with backpulse
- Decreased system complexity
- Increased permeate quality
- Reduced membrane wear

The pilot study utilized both relaxation and backpulsing during operation of the MBR pilot system. The relax frequency and duration remained constant at 10 minutes and 30 seconds, respectively. Backpulsing was utilized sporadically.

Figure 4-3: Permeability and temperature.



Air Scouring

Air scouring is another component of the cleaning process. Air is supplied to the bottom of the membrane module via an integrated coarse bubble aerator. As air bubbles travel to the surface of the tank, the outside of the membrane fibers are scoured, and any larger particles that may have adhered to the surface of the fibers are removed. Aeration is also used to sustain a minimum dissolved oxygen (DO) concentration of 2 mg/L in the tank, which is necessary to maintain a healthy bacterial population.

In this pilot study, the airflow in the tanks was initially 16 cfm, cycling in intervals at 10 seconds on and 10 seconds off. On July 25, the airflow increased to 30 cfm. Over the course of time, the efficiency of the blower declined, causing the airflow to decrease. On November 1, the airflow to the membrane module was recorded at 10 cfm. To improve the airflow to the tank, a second blower was installed and the airflow increased to 25 cfm. The cycling frequency of 10 seconds on and 10 seconds off was maintained throughout the study.

Figure 4-4 shows the DO concentration profile in the anoxic and aerobic tanks during the pilot study. A DO concentration greater than 1.5 mg/L is desired in the aerobic tanks for BOD_5 removal and nitrification. A DO less than 0.5 mg/L is desired in the anoxic tanks for denitrification. Prior to a change in the configuration of the aerobic and anoxic tanks, the dissolved oxygen (DO) concentrations in the anoxic and aerobic tanks were not on target. After November 7, the DO concentration in the aerobic tanks was generally higher than 1 mg/L and in the anoxic tanks, it was generally less than 0.2 mg/L.

ANALYTICAL PARAMETERS

Analytical results have been compiled (see Appendix A for a tabular listing of the data) and are plotted in Figures 4-5 to 4-11. Analytical parameters were measured by Suffolk County staff and by an independent laboratory. Both sets of results are presented, however the results from the lab are considered more accurate.

Mixed Liquor Suspended Solids (MLSS)

Figure 4-5 shows MLSS concentration in the Membrane Tank over the course of the study. The MBR system is designed to operate with a MLSS in the range of 8,000 to 12,000 mg/L, with a target MLSS of 10,000 mg/L. During the pilot study the MLSS as measured by the site ranged from 2,100 to 27,000 mg/L, with an average concentration of 8,065 mg/L. The laboratory results ranged from 190 to 12,320 mg/L, with an average MLSS concentration of 6,400 mg/L.

Figure 4-4: Dissolved Oxygen.

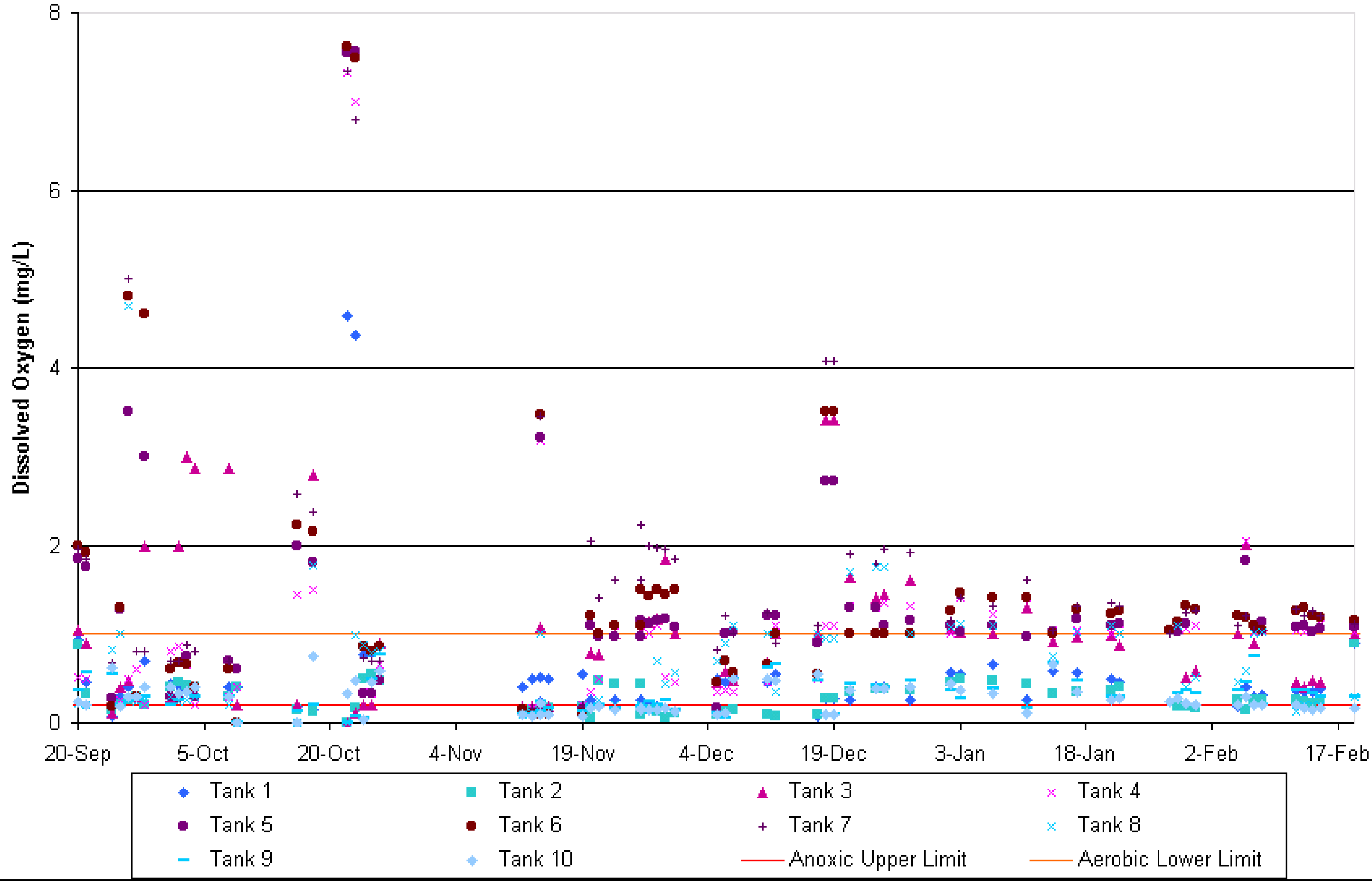
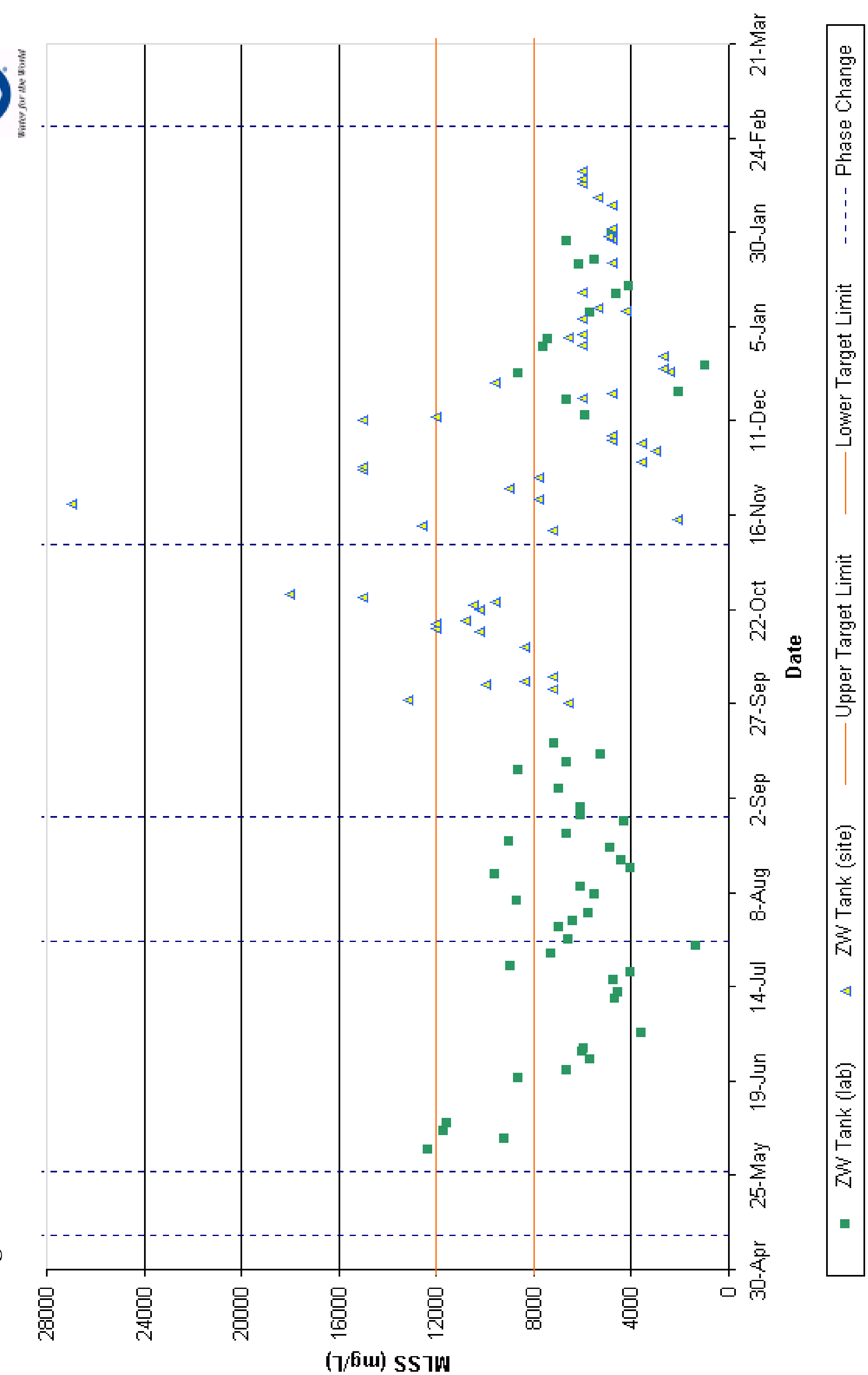


Figure 4-5: ZW Tank MLSS.



Nitrogen Species

Nitrogen in any soluble form is a nutrient and may need to be removed from wastewater to help control algae growth in the receiving body. Wastewater treatment facilities, which discharge treated effluent to the ground (subsurface discharge), may need to remove nitrogen in any soluble form (nitrate in particular) to minimize possible impact to acquifers. In addition, nitrogen in the form of ammonia exerts an oxygen demand and can be toxic to fish. Removal of nitrogen can be accomplished either biologically or chemically. The biological removal process of nitrogen species is called nitrification/denitrification. The nitrification/denitrification steps are expressed below:

1. Oxidation of ammonium to nitrite by Nitrosomonas microorganisms:

$$NH_4^+ + 1.5\ O_2 \rightarrow 2H^+ + H_2O + NO_2^-$$

2. Oxidation of nitrite to nitrate by Nitrobacter microorganisms:

$$NO_2^- + 0.5\ O_2 \rightarrow NO_3^-$$

The overall oxidation of ammonium, which is the nitrification step, is expressed below:

$$NH_4^+ + 2O_2 \rightarrow NO_3^- + 2H^+ + H_2O$$

3. The overall reduction of nitrate to nitrogen gas, the denitrification step, is expressed below:

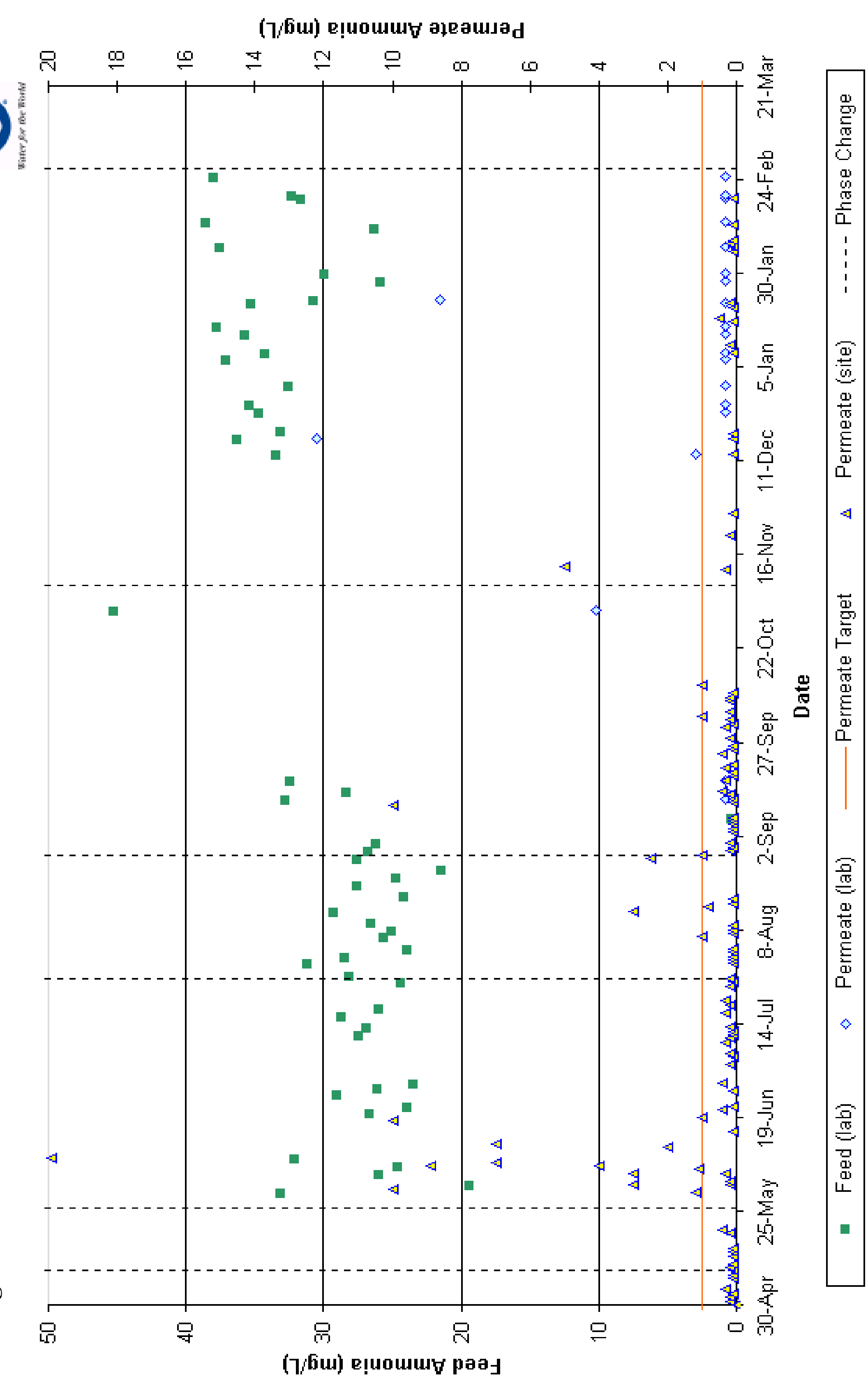
$$6NO_3^- + 5CH_nOH_m \rightarrow 5CO_2 + 7H_2O + 6OH^- + 3N_2$$

The CH_nOH_m represents carbonaceous BOD that the various denitrifying bacteria use as a carbon source. Where insufficient carbonaceous BOD is present for use as a carbon source, methanol addition is commonly practiced.

The degree of nitrification of wastewater is indicated by the relative amount of ammonia that is present. In an aerobic environment, bacteria can oxidize the ammonia-nitrogen to nitrites and nitrates. The predominance of nitrate-nitrogen in wastewater indicates that the waste has been stabilized with respect to oxygen demand.

Figure 4-6 shows the ammonia-nitrogen levels in the feed and permeate. Feed ammonia-nitrogen was measured between 19 and 45 mg/L. Based on results from the site, permeate ammonia-nitrogen ranged from 0.01 to 19.9, averaging 1.0 mg/L. After optimizing for nitrogen removal, 95% of the data points collected showed ammonia-nitrogen less than 1.0 mg/L in the permeate, which is indicative of near complete biological nitrification.

Figure 4-6: Ammonia.



Feed and permeate nitrite/nitrate levels are shown in Figure 4-7. Nitrites are short lived intermediate species that will not accumulate in a healthy nitrification system. Feed nitrates ranged from 0.1 to 11.5 mg/L based on lab results. Permeate nitrate levels recorded on site fluctuated from 0.1 to 20 mg/L. High nitrate concentrations were seen at the end of the study, when BOD_5 levels in the permeate were also high. It is thought that a number of shutdowns resulted in poor microorganism health, which in turn affected the denitrification step of the process.

Figure 4-8 shows the Total Kjeldahl Nitrogen (TKN) levels in the permeate, measured both at the lab and on site. At the beginning of the study, the TKN measured by the site ranged from 0.1 to 29 mg/L. However, from September to the end of the study, the permeate TKN was consistently less than 1.5 mg/L, as measured by the lab.

Figure 4-9 shows the total nitrogen concentration in the feed and permeate. Total nitrogen (TN) in the feed was calculated by adding the TKN value with nitrate and nitrite values as measured by the lab. TN in the permeate was calculated by adding the TKN value with the NO_x values, again as measured by the lab. Total nitrogen values greater than 50 mg/L in the permeate were considered erroneous, since the influent TKN was consistently less than 50 mg/L. After removing these values, the permeate TN ranged from 4.8 to 35.3 mg/L, with an average of 14.0 mg/L. During the period of December 24 – 31, 2001, when the pilot was running at the optimum conditions, the permeate TN ranged from 4.8 to 6.1 mg/L, with an average of 5.4 mg/L. These results were used to determine the lowest total nitrogen levels in the permeate achievable without methanol addition and also demonstrate that the no methanol addition is required to achieve a permeate TN level less than 8 mg/L in the permeate, when the system is running optimally. However, sustained operation while producing similar results is necessary before this process technology can be endorsed for this application.

As influent wastewater characteristic information was collected during the first portion of this study (5/30/01 – 7/25/01), the BOD:TKN ratio was found to be approximately 6.0. A BOD:TKN ratio of 4.0 or more is considered an acceptable range for nitrogen removal. Weaker wastewater (BOD:TKN < 4) typically requires methanol or other supplemental carbon sources to produce low (<3 mg/L) effluent TN concentrations. As such, methanol addition was thought to be unnecessary for remaining pilot activities.

BOD_5

Biochemical oxygen demand is a measurement of the amount of DO required to meet the metabolic needs of the microorganisms in order to degrade the organic matter in wastewater. Figure 4-10 shows the BOD_5 profile. During the first few months of the study, permeate BOD_5 levels less than 5 mg/L were consistently achieved. From November 2001 to February 2002, the permeate BOD_5 concentration was much more

Figure 4-7: Nitrates and Nitrites.

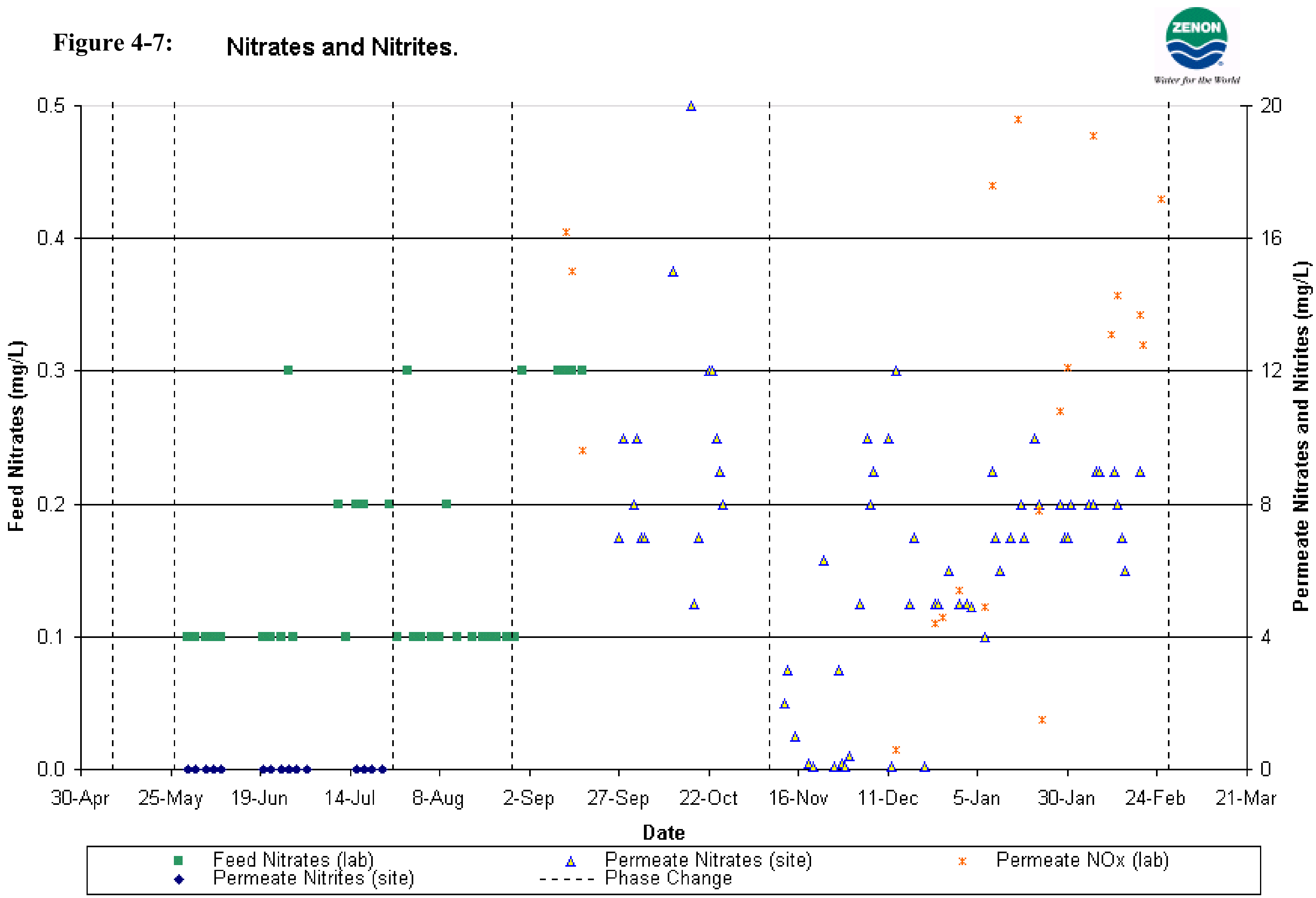


Figure 4-8: TKN.

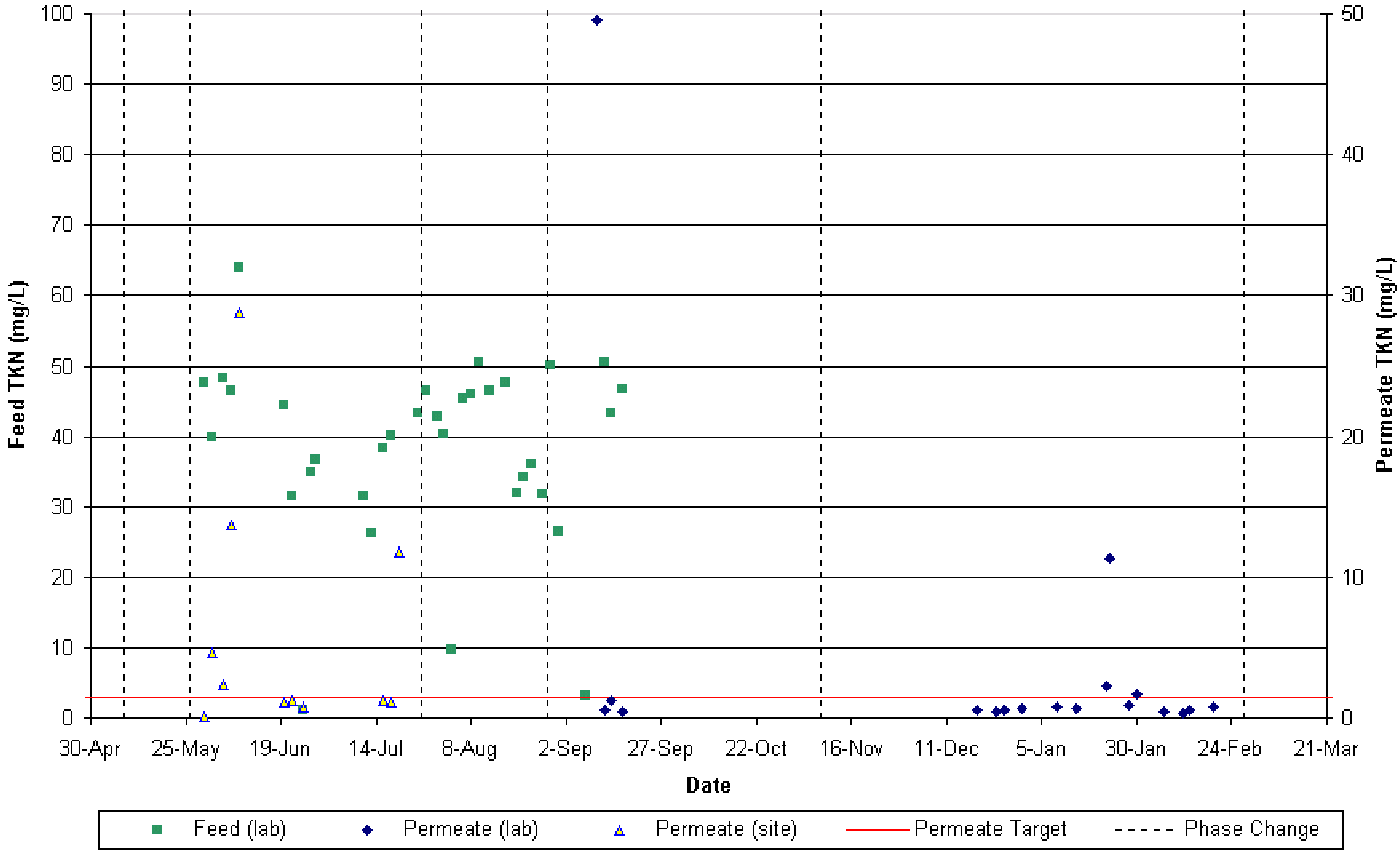


Figure 4-9: **Total Nitrogen.**

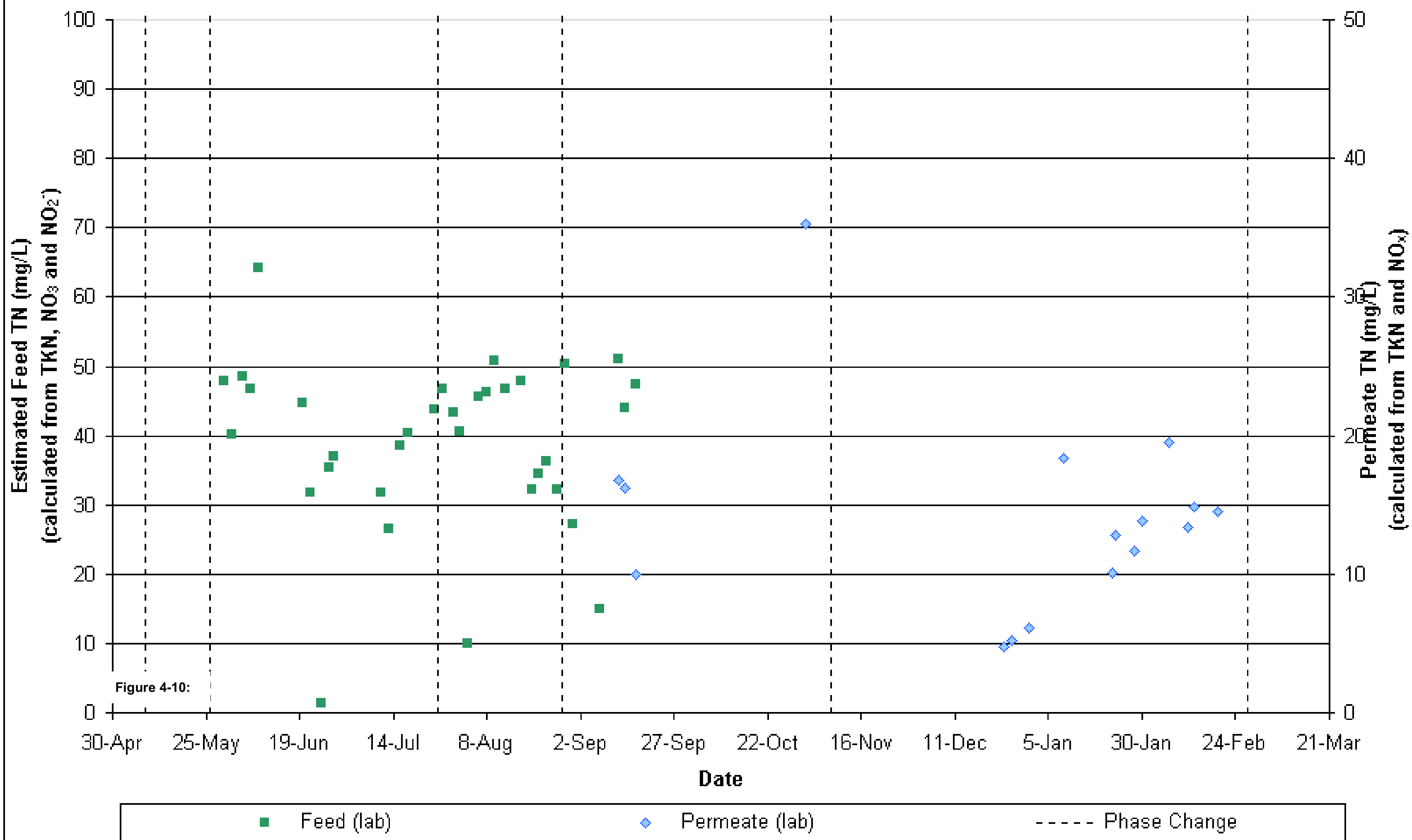
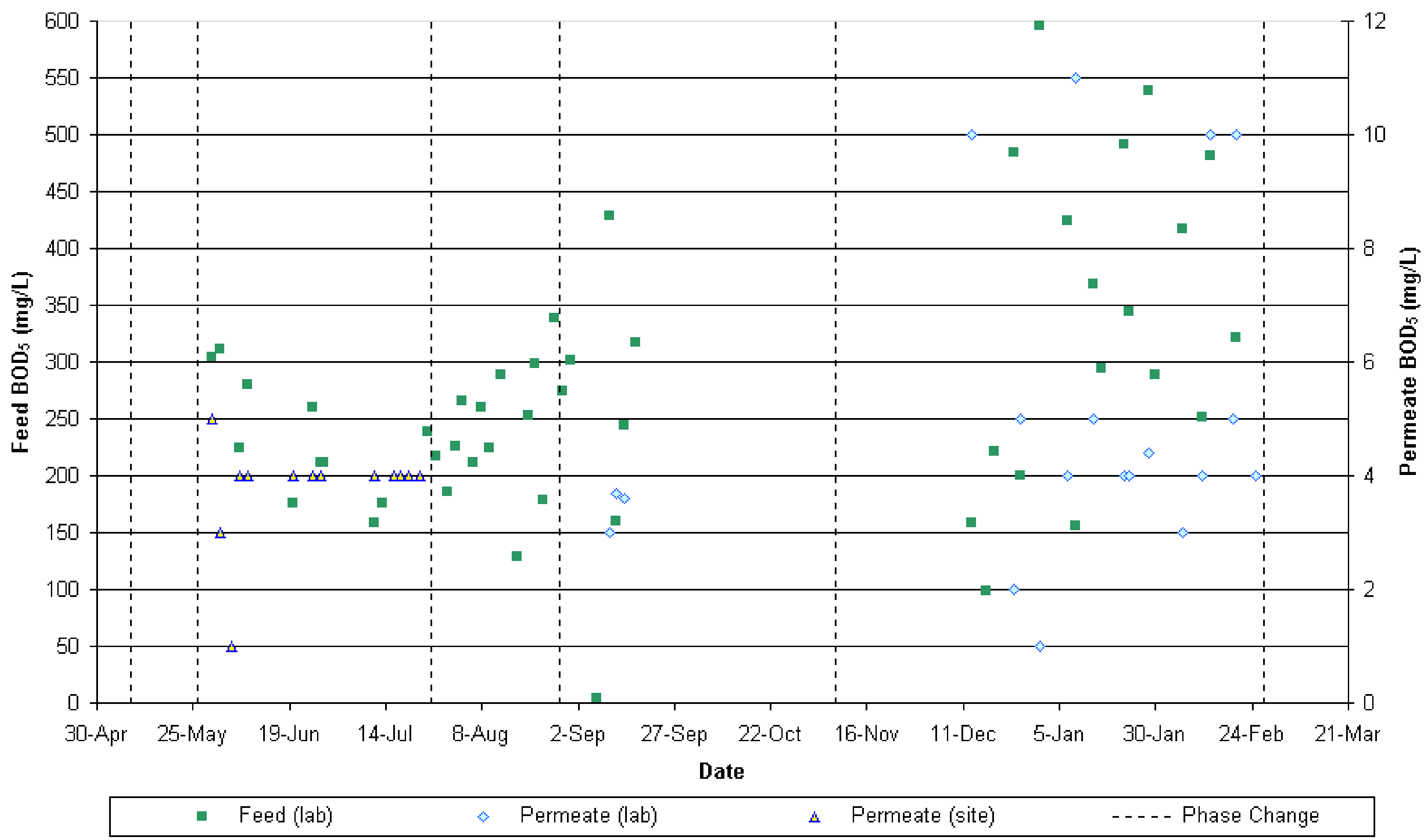


Figure 4-10: **BOD_5.**



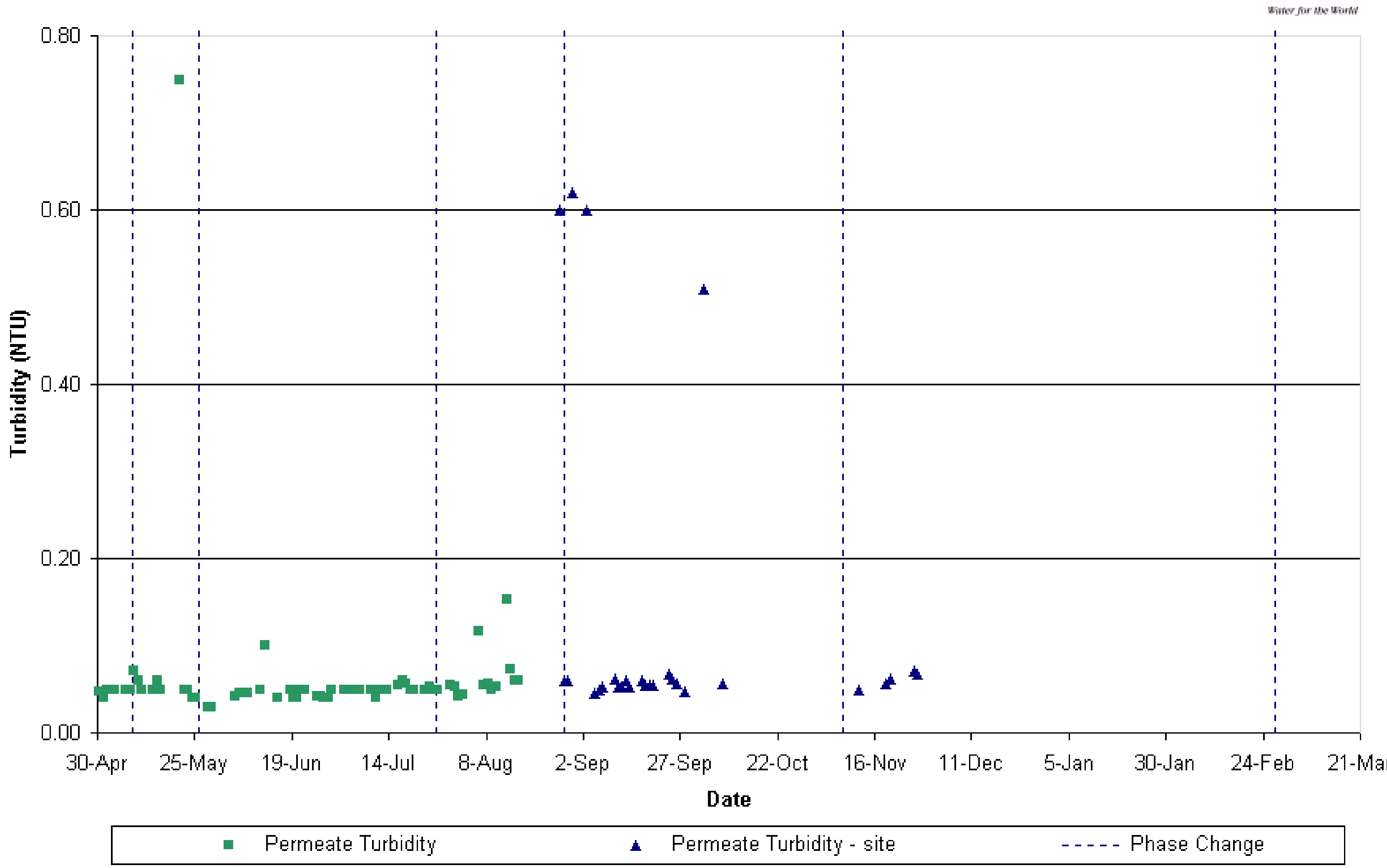
sporadic, ranging from 1 to 11 mg/L. These BOD_5 levels are indicative of poor microorganism health in the latter portion of the study, likely due in part to the number of shut downs experienced during this time.

Turbidity

Turbidity is a measure of the clarity of water and is commonly expressed in nephelometric turbidity units (NTU). Suspended solids and colloidal matter, such as clay, silt and microscopic organisms cause turbidity.

The MBR permeate turbidity is shown in Figure 4-11. Turbidity was not recorded after November 27, therefore this data is not included. Permeate turbidity remained close to 0.05 NTU for most of the study. A few measurements exceeded 0.1 NTU, likely due to fluctuations of flow to the turbidimeter and system shutdowns.

Figure 4-11: Permeate turbidity.



Section 5

PILOT OPERATION

A field testing and monitoring program was developed to achieve the objectives of the performance evaluation. The program consisted of a start-up phase and was planned to have four operational phases. The goal of all operational phases was to achieve $CBOD_5 < 5$ mg/L and TSS < 1 mg/L while measuring the amount of TN in the treated effluent. For Phase I, the goal was to determine the lowest achievable TN without methanol addition. The goal of Phase II was to determine the lowest achievable TN with methanol addition. The goal of Phase III was to determine the lowest methanol concentration necessary to achieve < 8 mg/L of TN. The goal of Phase IV was to measure performance under cold weather conditions. Phases II, III, and IV were not completed due to difficulties with the sustained operation of the pilot system and the length of time it took to complete Phase 1. This section discusses the results of the Phase I activities. At the end of the demonstration, membrane integrity was tested.

PHASE 1 – LOWEST TOTAL NITROGEN WITHOUT METHANOL

The field operation (Phase I) can be broken into five periods corresponding to changes in the pilot system operational set points and flow patterns that were made to achieve the best total nitrogen reduction performance. The key parameters varied during the periods are listed in Table 5-1 below.

Table 5-1 Phase 1 – Key Parameters

Parameter	*Period 1*	*Period 2*	*Period 3*	*Period 4*	*Period 5*
Dates	5/8/01 – 5/25/01	5/25/01 – 7/25/01	7/25/01 – 8/26/01	8/26/01 – 11/7/01	11/7/01 – 2/27/02
Instantaneous Flux (GFD)	11	11	11	11	11
Membrane Air Flow (cfm)	16	16	25	15	25
Maintenance Clean Frequency (#/week)	1	1	1	1-3	3
Recirculation Rate (gpm)	15	25	25	25	25
Layout	Figure 3-2	Figure 3-2	Figure 3-2	Figure 3-4	Figure 3-6
Process Flow	Figure 3-3	Figure 3-3	Figure 3-3	Figure 3-5	Figure 3-7
Methanol Addition	None	None	None	None	None

During Period 1, the initial set points for operation of the MBR pilot system were established. The transition to Period 2 was made when the recirculation rate was increased to 25 gpm. At the start of Period 3, the air flow to the membranes was increased to better maintain the permeate flux rate. For Periods 4 and

5, the process flow configuration was changed by altering the number of tanks operating in aerobic mode and changing the flow routing of the recirculation loops.

The operating data, based on samples collected at the site by Suffolk County staff and analyzed in a County operated laboratory, is included in Appendices A-1, A-2, A-3 and A-4. An operating event log for the Phase I pilot activities is included in Appendix B.

INITIAL START UP, SYSTEM SEEDING, AND ACCLIMATION (APRIL 10 TO MAY 8, 2001)

During initial start up, the pilot system was seeded with sludge from the Twelve Pines Sewage Treatment Plant. For the first month, the pilot unit was operated in a modified batch mode in order to increase the MLSS concentration in the Membrane Tank to the target level of 8,000 mg/L. Operational issues related to the equipment and the methods used for analytical sampling delayed the acclimation of the pilot system. On May 8, a MLSS concentration of 8,000 mg/L in the Membrane Tank was achieved and the pilot operation began.

PERIOD 1: DIRECT FILTRATION (MAY 9 TO MAY 25, 2001)

Period 1 is the time when plant staff became acquainted with the continuous operation of the pilot system, alarm set points were fine tuned and sample collection procedures were established. Daily samples were not collected during this period, sampling was done sporadically to check the pilot system performance.

During this period, the permeate flux rate was set at 11 GFD and a relax frequency of 10 minutes for a duration of 30 seconds was used. Maintenance cleaning of the membranes was done once each week with sodium hypochlorite at a concentration of 200 mg/L. The air to the membranes was set at 16 cfm with on/off cycles set to 10 seconds. The system vacuum pressure was very stable at 1 psi during this period.

Reported measurements for MLSS showed the concentration in the Membrane Tank increased from 8,100 mg/L up to 24,000 mg/L. The validity of these results is questionable, due to the inconsistent trend in the numbers.

Ammonia-nitrogen was measured by site personnel during this period. Results showed that ammonia-nitrogen levels in the permeate ranged from 0.1 to 0.4 mg/L. Permeate turbidity was less than 0.07 NTU 97% of the time.

PERIOD 2: INCREASED RECIRCULATION RATES (MAY 26 TO JULY 25, 2001)

On May 25, the recirculation flow from Tank 10 to the Membrane Tank was increased to 25 gpm from 15 gpm to improve the mixing in these tanks by "turning them over" more frequently. The flux remained at 11 GFD and the relax frequency/duration was maintained at 10 minutes and 30 seconds respectively. During this period, the vacuum increased as high as 2.5 psi, but was generally stable at 0.5 psi. All other operational parameters remained the same. The operating data from this period is listed in Appendix A-1.

At the beginning of Period 2, the MLSS concentration in the Membrane Tank was quite high, ranging from 8,640 to 15,600 mg/L with one outlier at 26,400 mg/L. The MLSS concentration decreased to between 3,000 and 6,000 mg/L around June 19 and remained close to this level for the rest of the period. Since no sludge was wasted during Period 2, this decrease in MLSS was unexpected. A likely explanation for this anomaly is that the solids were accumulating in the anoxic tanks, which lacked sufficient mixing at that time. The presence of thick sludge blankets in these tanks was later observed when there was insufficient mixing.

Despite the mechanical problems experienced at the beginning of the period and the resultant system shutdowns, analytical parameters were measured by site staff. Permeate ammonia-nitrogen and TKN levels were high during these few weeks. Ammonia-nitrogen did drop to between 0.1 and 0.4 mg/L and TKN dropped below 1.5 mg/L by June 19, correlating to the drop in MLSS concentration. This correlation was likely the result of too little oxygen supplied when the solids inventory in the system was high, which limited the ability of the microbes to perform nitrification. Permeate BOD_5 was fairly stable at 4 mg/L during this period, while permeate turbidity was very good at less than 0.1 NTU, 100% of the time.

Late in the period, black sludge and a strong smell was observed in the aerobic tanks. At the same time, the MLSS concentration increased rapidly from approximately 4,000 mg/L to 9,000 mg/L. It is likely that a portion of the anaerobic sludge blanket that had been amassing in the anoxic zones was recirculated into the system, disrupting the balance of the microbial population in the aerobic zones. To restabilize the mixed liquor, approximately 1,500 gallons of sludge was wasted on July 24.

Operating data for this period is summarized herein:

- Average effluent BOD_5 was 3.79 mg/L, with 100% of the values at <5 mg/L
- Average effluent TSS was 3.47 mg/L, with 48% of values <1 mg/L
- Average effluent NH_3 was 2.26 mg/L
- Average effluent TKN was 4.32 mg/L
- Average effluent NO_3 was 9.32 mg/L
- Average effluent NO_2 was 0.1 mg/L

- Average effluent TN was 13.6 mg/L, with the lowest measured value of 0.8 mg/L
- Average effluent TKN was 4.32 mg/L
- Average effluent turbidity was <0.1 NTU

PERIOD 3: INCREASED AIR TO MEMBRANES (JULY 25 TO AUGUST 27, 2001)

Over the course of Periods 1 and 2, the aeration to the membranes was set to 16 cfm. During Period 2, the efficiency of the blower started to decline and an additional blower was sent to the site to supplement the airflow to the membrane. Installation of this blower occurred late in July. The membrane system was returned to service with airflow to the membrane increased to 25 cfm, in cycles of 10 seconds. The flux was maintained at 11 GFD, and the recirculation rates of 15 gpm and 25 gpm were kept constant for the duration of the Period 3. The operating data form this period is listed in Appendix A-2.

The MLSS concentration recorded on July 25 was very low, measured at 1,340 mg/L. This result is likely due to the wasting half of the system inventory towards the end of Period 2. For the rest of the period, the MLSS concentration was between 4,000 and 9,560 mg/L, with most samples falling under the targeted concentration of 8,000 mg/L.

Permeate ammonia-nitrogen results during this period were very good, however one sample was recorded at 3 mg/L on August 13, but all other samples fell below the target of 1 mg/L. Only two measurements of turbidity in the permeate exceeded 0.1 NTU.

Operating data for this period included:

- Average effluent BOD5 was 7.25 mg/L, with 77% of the values at <5 mg/L
- Average effluent TSS was 2.14 mg/L, with 79% of values <1 mg/L
- Average effluent TN was 17.2 mg/L, with the lowest value achieved 10.6 mg/L
- Average effluent NH3 was 0.1 mg/L
- Average effluent TKN was 3.61 mg/L
- Average effluent NOx was 32.1 mg/L
- Average effluent turbidity was <0.1 NTU

PERIOD 4: CHANGE IN TANK CONFIGURATION (AUGUST 27 TO NOVEMBER 7, 2001)

After analysis of the results of Periods 1 through 3, a decision was made to change the configuration of the tanks (Figures 3-4 and 3-5) to improve nitrogen removal. On August 27, aeration to Tank 8 was ceased, and the tank was converted to an anoxic operation. In the original process scheme, the overflow from the Membrane Tank was directed to Tank 1, resulting in high concentrations of DO in the first anoxic tank and negatively impacting the denitrification in this zone. On August 27, the overflow was diverted to Tank 3, an aerobic tank. Later in the study, it was determined that the diversion had not been properly completed, as two lines had connected the Membrane Tank to Tank 1, and only one had been moved to Tank 3. On November 7, this was rectified, and the entire overflow was diverted to Tank 3. The operating data for this period is shown in Appendix A-3.

Flux during this period was maintained at 11 GFD, and the recirculation rates at 15 gpm and 25 gpm for the inner and outer loops, respectively. Mechanical problems were experienced with the supplemental blower, which was taken off-line during this period, resulting in a decreased airflow to the membrane of 15 cfm.

For the first three weeks of this period, the vacuum was very constant around 1 psi. On September 26, the vacuum increased to 2 psi, and continued to climb over the next 9 days ultimately reaching 4.4 psi. For the first few weeks of October, the vacuum remained high and the operators performed daily maintenance cleans with sodium hypochlorite to reduce the vacuum. During the last two weeks of October, the MBR system continued to operate at a high vacuum, experiencing several alarms. After aerating the membrane overnight, the vacuum dropped from 10 psi to 1.5 psi without the need for a chemical recovery clean. The operation of the system throughout October was not consistent, resulting in less meaningful analytical data.

On October 31, a ZENON representative arrived at the site to determine the cause of the high vacuum situation. The conditions of the pilot unit were also checked at this time, and found to be off-target. Table 5-2, presents the target and actual values of the system parameters on November 1.

Table 5-2 MBR Pilot Key Operating Parameters Target vs. Actual Conditions as of Nov. 1, 2001

Parameter	*Target*	*Actual*
Flux (GFD)	11	11
Permeate and Relax duration (min/sec)	10/30	10/30
Recirculation pump #1 (gpm)	15	1
Recirculation pump #2 (gpm)	25	30
Membrane Tank aeration (cfm)	25	10
Aerobic tank aeration (cfm)	6	2

It was also discovered that the mixed liquor overflow from the MBR tank had not been properly diverted from Tank 1 to Tank 3, as mentioned earlier.

The MLSS concentration in the Membrane Tank started out low at the beginning of Period 4, but reached the target of 8,000 mg/L by September 5. The concentration then fluctuated between 6,000 and 18,000 mg/L for the remainder of the period.

Permeate ammonia-nitrogen levels measured at site during this period were excellent, falling below the target of 1 mg/L, 94% of the time, and below 0.5 mg/L, 85% of the time. Only a few BOD5 samples were collected and the results indicated a permeate BOD5 concentration of 3 to 4 mg/L.

Operating data for this period were:

- Average effluent BOD5 was 3.6 mg/L, with 100% of the values at <5 mg/L
- Average effluent TSS was 3.3 mg/L, with 25% of values <1 mg/L
- Average effluent TN was 36.1 mg/L, with the lowest value achieved 9.6 mg/L
- Average effluent NH3 was 7.02 mg/L
- Average effluent TKN was 13 mg/L
- Average effluent NOx was 23.1 mg/L

PERIOD 5: CHANGE IN TANK CONFIGURATION II (NOVEMBER 7 TO FEBRUARY 27, 2002)

In addition to re-establishing the desired parameters of the pilot (Table 5-2), several other mechanical issues were resolved before Period 5 was started. The bag filter housing in the feed line to the pilot was unclogged and the sampling ports on each tank were also cleared of debris.

Mixing of the anoxic tanks was also addressed. Until this point, mixing in the anoxic zones was minimal. In October, valves had been installed in the anoxic zone, which would allow a 10 second pulse of air into Tanks 2, 8, 9 and 10 every 20 minutes, to aid in the mixing of the contents of these tanks. While on site, ZENON's representative discovered that the first anoxic tank (Tank 1) was still not being mixed as the aeration grid had not been installed. To keep the tank properly mixed, a submersible pump was installed to continuously agitate the contents of the tank.

On November 7, the MBR pilot system was restarted at 11 GFD flux, recirculation rates of 25 and 15 gpm for the outer and inner loops, respectively, and aeration to the membrane at 25 cfm. Mixing in the anoxic zones was obtained using pulses of air for 10 seconds every 20 minutes, and air was introduced to the aerobic zones at 6 cfm. Maintenance cleaning was not conducted at the beginning of this period.

For most of the month of November, the vacuum remained around 1 to 2 psi. At the end of November, the vacuum increased, causing a high level alarm. It was later determined that the increase in pressure was a result of blower failure causing a lack of air to the membranes.

For the rest of this period, multiple shutdowns were experienced for a variety of reasons that can be noted in the Event Log included as Appendix B.

One time late in the period to address a high vacuum alarm, a maintenance cleaning was conducted on the membrane with approximately 500 mg/L of chlorine. The cleaning consisted of backpulsing and relaxing the membrane for 60 and 300 seconds respectively. This routine was conducted 10 times. The membrane was allowed to soak overnight in chlorine. This procedure, however, did not result in a substantially lower vacuum, and therefore a recovery clean was started.

MLSS levels ranging between 2,100 mg/L and 27,000 mg/L were recorded during November and December, however most MLSS measurements made during Period 5 were recorded between 4,000 and 7,000 mg/L. At times when the MLSS concentration was low, the nitrate results were slightly higher. Throughout February, the readings for the MLSS concentration in the Membrane Tank were low. On February 6, the concentration was measured at 4,800 mg/L. By February 13, the concentration had increased to 6,000 mg/L and remained there until February 20.

During this period, the permeate ammonia-nitrogen concentration measured at site was below 0.3 mg/L, 85% of the time, and was below 1 mg/L, 95% of the time. On November 13, December 17 and January 23, high permeate ammonia-nitrogen concentrations were recorded. These increases can be attributed to loss of air to the aerobic tanks due to power failure.

During the month of December, when the system was operating consistently, low total nitrogen levels were seen in the permeate. The TN ranged from 4.8 to 6.1 mg/L with an average of 5.4 mg/L.

The permeate BOD5 concentration during Period 5 ranged between 1 and 11 mg/L. A BOD5 concentration greater than 5 mg/L in the permeate generally indicates problems with the process. In this instance, a number of factors could have contributed to the high BOD5 levels, including temperature variances, low MLSS concentrations, process shut downs resulting in disturbances of the microorganism population and possible algal and other organic contamination. The sludge blanket seen in several tanks likely contributed to the poor BOD5 results recorded during this period because of the reduced working volume of the system and poor circulation of the tank contents.

Operating data for this period is included in Appendix A-4 and is summarized herein:

- Average effluent BOD5 was 5.4 mg/L, with 75% of the values at <5 mg/L
- Average effluent TSS was 3.2 mg/L, with 42% of values <1 mg/L

- Average effluent TN was 20.6 mg/L, with the lowest value achieved 4.8 mg/L
- Average effluent NH_3 was 0.88 mg/L
- Average effluent TKN was 1.6 mg/L
- Average effluent NO_x was 16.6 mg/L

PHASES 2, 3 & 4

The additional planned phases were not completed due to the length of time it took to get reliable operation to complete the first phase of the program. However, information from other pilot and full scale MBR systems was gathered to show the performance of this technology under the operating conditions planned for the subsequent phases of the test program. This information is discussed in Section 7 of this report.

MEMBRANE INTEGRITY

Prior to the start up of the study, tests were conducted on the membrane fibers, including tests for tensile strength and molecular weight cut-off. The tensile strength of the individual fibers is greater than 100 pounds.

A membrane integrity test was performed during the start up of the pilot study via bubble-point observation. Results of this test were positive with no discernable bubble streams detected when the membrane was pressurized up to 5 psi.

Tests were also conducted to determine the membrane permeability prior to the study. Clean membrane permeability was measured at 14.1 GFD/psi at 20oC.

Permeate turbidity was monitored throughout the study, though not recorded after November 27. The data collected shows no breach of membrane integrity, as 96% of the measurements showed turbidity less than 0.1 NTU. Data recorded above 0.1 NTU was likely due to system shut downs or fluctuations in the flow to the turbidimeter.

At the end of the study, the membrane was cleaned and the permeability was measured to be 22.2 GFD/psi. The higher permeability recorded at the end of the study was likely due to the imprecise measurements of low vacuum conditions. For example, a vacuum reading of 0.7 psi at 10 GFD flux and 20oC corresponds to a membrane permeability of 14.3 GFD/psi. A vacuum reading of 0.5 psi at 10 GFD flux and 20oC corresponds to a membrane permeability of 20 GFD/psi. Therefore, under these membrane conditions, a difference of 0.2 psi results in a large difference in membrane permeability.

Upon return of the pilot equipment to the ZENON factory, further tests were conducted on the membrane fibers. There was no discernable difference between the fibers used in the Suffolk County test and new fibers in terms of tensile strength and molecular weight cut off.

CLEANING

Two types of membrane cleaning techniques are employed at full-scale municipal MBR facilities. The first type is maintenance cleaning. The membranes are not removed from the aeration tank for this type of cleaning. In the full-scale systems, the procedure is entirely automated and scheduled to occur during off-peak hours of the day when the membranes would otherwise be in standby mode. The procedure is an extended backpulse conducted over a one-hour period. Approximately 200 mg/L of sodium hypochlorite, or 2,000 mg/L of citric acid, is backpulsed through the membranes at regular intervals over the one-hour period. The procedure is normally conducted three to seven times per week.

In this study, maintenance cleaning was conducted with 200 mg/L of sodium hypochlorite. At the beginning of the study, this type of cleaning was initiated on a weekly basis. Later, maintenance cleaning was performed three times a week, or as required. During Periods 4 and 5, when a number of high vacuum alarms were experienced, maintenance cleaning was conducted on a daily basis.

The second type of cleaning is termed recovery cleaning. Individual membrane cassettes are removed from the aeration tank and sprayed down to remove accumulated mixed liquor solids. The membrane cassette is transported to a separate membrane-soaking tank and immersed for a twelve-hour period in 1000 mg/L of sodium hypochlorite (or 2,000 mg/L citric acid). Individual cassettes are cleansed at intervals ranging from once every 3 months to once per year.

A recovery cleaning is required to restore the permeability of the membrane once the membrane becomes fouled. A recovery cleaning should be initiated when permeability declines to less than 50% of initial stable permeability. This will generally occur when the vacuum exceeds 9 psi. The cleaning chemicals that are typically used are sodium hypochlorite (NaOCl), for the removal of organic foulants, and citric acid, for the removal of inorganic contaminants.

One recovery cleaning was performed at the end of this pilot study. The cleaning was started by backpulsing 2000 mg/L of sodium hypochlorite through the membrane, then allowing the membrane to soak overnight at 200 mg/L. After this seven-hour soak, the membrane vacuum was still quite high, so a citric acid clean was conducted. Citric acid was backpulsed through the membrane at 10,000 mg/L, and the membrane was allowed to soak for several days in a solution of 1,000 mg/L citric acid. Once the system was restarted, the vacuum was less than 1 psi. It is likely that the addition of chlorine during the first portion of the cleaning elevated the pH in the Membrane Tank, causing scaling of the membrane. With the pH lowered during the citric acid clean, the scaling was easily removed and the membrane permeability restored.

Section 6
ECONOMIC ANALYSIS

MBR SYSTEM ESTIMATED COST

Based on data generated during the pilot, information gathered from MBR system suppliers, and published literature, capital, operating, and maintenance costs were estimated. The estimates are based on a system having capacity of 300,000 gpd average daily flow and achieving an effluent quality of $CBOD_5$ <5 mg/L, TSS <1 mg/L and ammonia-nitrogen <1 mg/L. Since it appeared that methanol addition would be necessary to achieve the targeted total nitrogen removal performance (TN <8 mg/L), the economic analysis assumed this consumption would be similar for both treatment systems and therefore, costs associated with methanol addition were not included in the analysis. The cost estimate is summarized in Table 6-1.

Table 6-1 MBR System Cost Estimate[1]

Description	*Cost*
Estimated Capital Cost	
• site and civil work	$15,000
• process equipment	$1,180,000
• process tank	$130,000
• process piping, valves, fittings	$35,000
• electrical, instrumentation, control	$135,000
subtotal	$1,495,000
engineering, legal, misc (25%)	$374,000
Estimated MBR System Capital Cost	$1,869,000
Estimated Annual Operating and Maintenance Costs	
• power[2]	$39,300/yr
• parts and repairs[3]	$15,000/yr
• chemicals[3]	$2,000/yr
• manufacturer service (routine and annual)[3]	$12,000/yr
• operations[4]	$37,400/yr
Estimated MBR System Operating Cost	$105,700/yr

(1) Based on 0.3 MGD average daily flow capacity system with a 0.6 MGD daily peak.
(2) Based on 327,500 kw-hrs/yr at $0.12/kw-hr.
(3) MBR system manufacturers recommendation.
(4) Based on 16 hrs/wk at $45/hr.

CONVENTIONAL ACTIVIATED SLUDGE SYSTEM ESTIMATED COSTS

A cost estimate for a conventional activated sludge process (sequencing batch reactor (SBR) technology) with tertiary filters was also prepared based on information from SBR and filter systems suppliers. The design capacity of the system is 300,000 gpd average daily flow capacity system and achieving an effluent quality of $CBOD_5$ <5 mg/L, TSS <1 mg/L and ammonia-nitrogen <1 mg/L. For comparison purposes, it has also been assumed that the total nitrogen removal with this technology can be achieved methanol addition.

The cost estimate is summarized in Table 6-2.

Table 6-2 Activated Sludge (SBR) System Cost Estimate [(1)]

Description	*Cost*
Estimated Capital Cost	
• site and civil work	$70,000
• process equipment (SBR)	$360,000
• process equipment (filters)	$260,000
• process tanks	$445,000
• process piping, valves, fittings	$85,000
• electrical, instrumentation, control	$120,000
subtotal	$1,340,000
engineering, legal, misc (25%)	$335,000
Estimated SBR System Capital Cost	$1,675,000
Estimated Annual Operating and Maintenance Costs	
• power[(2)]	$29,500/yr
• parts and repairs[(3)]	$9,300/yr
• chemicals[(4)]	-----
• manufacturer service (routine and annual)[(5)]	-----
• operations[(6)]	$37,400/yr
Estimated SBR System Operating Cost	$76,200/yr

(1) Based on 0.3 MGD average daily flow capacity system with a 0.6 MGD daily peak.
(2) Based on 246,000 kw-hrs/yr at $0.12/kw-hr.
(3) Based on 1.5% of equipment cost.
(4) None required.
(5) None required.
(6) Based on 16 hrs/wk at $45/hr.

The economic comparison of the two treatment systems is shown in Table 6-3.

Table 6-3 Economic Comparison MBR System and Convention System [1]

	MBR System	***Conventional System***
Estimated Capital Cost	$1,900,000	$1,700,000
Estimated Annual O&M Costs	$105,700	$76,200
Total Present Worth of Capital and O&M Costs[1]	$3,336,500	$2,735,600
Total Annual Cost of Capital and O&M Costs [1]	$245,500	$201,300

[1] Based on 4% interest, 20 years

Section 7

MEMBRANE BIOREACTOR SYSTEM PERFORMANCE

A summary of the performance of the Twelve Pines MBR pilot operation is included in Table 7-1.

Table 7-1 *Twelve Pines WWTP MBR Pilot Operation Performance Summary*

	BOD_5 (mg/L)		TSS (mg/L)		NH_3 (mg/L)		TKN (mg/L)		NO_2 (mg/L)		NO_3 (mg/L)		TN (mg/L)	
	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.	Inf.	Eff.
Period 2														
Ave	248	3.8	250	3.5	27	2.3	42	4.3	0.1	---	0.1	9.3	42.2	13.6
Max	624	5.0	578	15	38	19.9	64	28.8	0.2	---	0.3	17.7	---	28.9
Period 3														
Ave	228	7.3	263	2.1	27	0.1	43	3.6	0.1	---	6.6	---	43	17.2
Max	340	39	382	11	31	0.1	52	27.7	0.2	---	14	---	---	28.7
Period 4														
Ave	288	3.6	230	3.3	44	7	---	13	---	---	---	---	---	33
Max	428	4.0	438	8	81	37	---	49.6	---	---	---	---	---	88.9
Period 5														
Ave	371	5.4	519	3.2	34	0.7	---	1.6	---	---	---	---	---	20.6
Max	662	11	1160	10	39	8.6	---	11.3	---	---	---	---	---	122

These data show that the pilot MBR operation was able to achieve BOD_5 effluent objectives of <5 mg/L as demonstrated during Periods 2 & 4. TSS in the treated effluent was quite low, however the objective of <1 mg/L was not achieved. The TN objective of <8 mg/L was achieved for short periods, but this performance was not sustained and the objectives were not consistently demonstrated. High levels of nitrification (effluent NH_3-N <0.5) were demonstrated especially during in Period 2.

MBR PERFORMANCE AT OTHER FACILITIES

As total nitrogen removal objectives were not achieved during the Twelve Pines MBR pilot demonstration, operating data from other selected pilot and full-scale facilities were reviewed. This information from the most pertinent facility is summarized herein.

BROAD RUN WATER RELCAIMATION FACILITY MBR PILOT TESTING, LOUDOUN COUNTY, VA

An on-site MBR pilot project was conducted at the Leesburg, VA Water Pollution Control Facility (WPCF) from October 2000 through May 2001. The MBR influent utilized primary effluent from the WPCF. The pilot project is described in a document entitled: "Final Report for the Broad Run Water Reclamation Facility Pilot Testing Program", Loudoun County Sanitation Authority, August 2001.

The MBR's operating conditions and effluent results are summarized in Table 7-2.

Table 7-2 MBR Pilot Summary Broad Run WRF

Biological Treatment Operating Conditions	***Target***
Process Configurations	• 4-Stage Process with a De-aeration Zone (Modified Ludzak-Ettinger (MLE) Recycle Flows) • 5-Stage Operation • 4-Stage Operation
Hydraulic Retention Time (HRT)	• 8.4 hours (Average) • 5.6 hours (Peak)
Solids Retention Time (SRT)	• 19 to 23 days (30 days during startup)
Typical DO (mg/L)	• Anaerobic and Anoxic Zones (Zones 1, 2, 3, 5) 0.0 – 0.2 mg/L • Aerobic Zone (Zone 4) 0.5 – 1.5 mg/L • Aerobic Zone (Zone 6) Not Specified

Membrane Operating Conditions	***Target***
Membrane Flux	• 20.4 GFD (average) • 30.6 GFD (diurnal peak)
Permeate Flow	• 14.2 gpm (average) • 21.3 gpm (peak)
Membrane Aeration Mode	• Intermittent (10 seconds ON and 10 seconds OFF per pair of membranes)
Backpulse Frequency	• 10 minutes
Backpulse Duration	• 30 seconds
Backpulse Chemical Addition	• 2 to 4 mg/L sodium hypochlorite
Backpulse Flow Rate	• 1.5 times average flow
Maintenance Cleaning	• 2 to 7 cleanings per week
Chemical Addition for Maintenance Cleaning	• 200 mg/L Cl_2 residual

Reported Effluent	
BOD_5 (mg/L)	<2.0
TSS (mg/L)	<1.0
TKN (mg/L)	1.3 average [(1)]
NH_3 (mg/L)	<1.0
TN (mg/L)	5.6 average [(2)]
TP (mg/L)	0.03 average [(2)]

[(1)] 5 stage reactor with approximately 73 mg/L methanol addition.
[(2)] With biological phosphorus removal and approximately 70 mg/L alum addition.

Section 8
CONCLUSIONS

The following conclusions can be drawn from this pilot study:

- MBR system effluent (permeate) ammonia-nitrogen levels less than 1 mg/L were readily achieved when proper process conditions were attained.

- Permeate BOD_5 levels were consistently less than the study goal of 5 mg/L when the system was operating within appropriate parameter ranges and healthy microorganisms were maintained.

- A recovery cleaning did not have to be conducted on the membranes until the system had been operated for nine months. Regular maintenance cleaning and proper aeration of the membranes resulted in a recovery cleaning interval greater than the normal, manufacturer recommended period of six months.

- Total nitrogen levels of less than 8 mg/L in the permeate were achievable for short periods, albeit not consistently, without chemical addition.

- Total nitrogen levels of less than 8 mg/L have been successfully achieved at other full scale MBR operating installations with the use of methanol for denitrification.

- An economic analysis indicates that MBR systems can cost approximately 10 to 15% more to construct and approximately 33% more to operate than conventional (SBR) biological treatment systems using effluent filtration.

- The inability of the pilot unit to attain proper and reliable process operating conditions during portions of this study due to a variety of reasons needs to be addressed before conducting further studies with this particular equipment.

REFERENCES

CH2MHILL. Broad Run WRF Pilot Testing Program Final Report. August 2001, pp 2-19, 8-1.

Hare, R.W., Sutton, P.M., Mishra, P.N. and A. Janson, "Membrane Enhanced Biological Treatment of Oily Wastewater," presented at the 63rd Annual Conference of the Water Pollution Control Federation, Washington, D.C., October 1990.

Metcalf & Eddy, Inc. Wastewater Engineering: Treatment, Disposal and Reuse. Revised by George Tchabanoglous and Franklin L. Burton. McGraw Hill Inc., 1991.

Reed, Sherwood C., Crites, Ronald W., and Middlebrooks, E. Joe. Natural Systems for Waste Management and Treatment, 2nd ed. McGraw Hill Inc., New York, 1995.

Thompson, D., Mourato, D., Penny, J., "Demonstration of the ZenoGem® Process for Municipal Wastewater Treatment," presented at the 71st WEFTEC Conference, Orlando, October 1998.

APPENDIX A

		LOCATION 1 (FEED)							LOCATION 2 (PERMEATE)									
						Field	Field	Field								Field	Field	Field
Day	Date	BOD_5	TSS	NH3	Alkalinity	Temperature	pH	Alkalinity	BOD_5	TSS	NH3	TKN	NOx	Total Nitrogen	Alkalinity	pH	Alkalinity	NO3
	(M/D/Y)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	°F		(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)			(mg/l)		(mg/l)	(mg/l)
F	11/9/2001																	
M	11/12/2001						7.5	240								7.1	80	2
T	11/13/2001						7.1	200								7	220	3
W	11/14/2001						N/A	N/A								N/A	N/A	N/A
T							N/A	N/A								7	220	1
F	11/16/2001																	
M	11/19/2001						7.1	180								6.9	120	0.2
T						64	6.9	200								6.9	100	0.1
W	11/21/2001	N/A	N/A	N/A	N/A				N/A	N/A	N/A	N/A	N/A	N/A	N/A			
T																		
F	11/23/2001					68	N/A	100								6.7	180	6.3
M	11/26/2001	N/A	N/A	N/A	N/A	66	6.9	260	N/A	N/A	N/A	N/A	N/A	N/A	N/A	6.9	100	0.1
T							N/A	N/A								N/A	N/A	3
W	11/28/2001	N/A	N/A	N/A	N/A		6.8	280	N/A	N/A	N/A	N/A	N/A	N/A	N/A	6.9	120	0.2
T							7	240								6.9	120	0.1
F	11/30/2001						7.2	160								7	100	0.4
	12/1/2001																	
M	12/3/2001	N/A	N/A	N/A	N/A		6.8	N/A	N/A	N/A	N/A	N/A	N/A	N/A	N/A	7.2	160	5
T																		
W	12/5/2001	N/A	N/A	N/A	N/A		7.3	220	N/A	N/A	N/A	N/A	N/A	N/A	N/A	6.8	160	10
T	12/6/2001						7	220								6.5	140	8
F	12/7/2001						7.2	160								7	120	9
M	12/10/2001																	
T	12/11/2001					65	7.3	240								6.5	180	10
W	12/12/2001					65	7.4	260								6.5	140	0.1
T	12/13/2001	159	76	33.5	N/A	66	7.2	200	10	4	1.2	N/A	0.6	N/A	N/A	6.7	180	6
F	12/14/2001																	
M	12/17/2001	99	576	36.3	N/A	69	6.8	160							N/A	6.7	140	5
T						69	6.8	240								6.7	120	7
W	12/19/2001	221	216	33.1	N/A				N/A	N/A	N/A	0.6	N/A	N/A	N/A			
T																		
F	12/21/2001					60	7.1	160								6.7	100	0.1
M	12/24/2001	485	660	34.7	N/A	68	6.6	N/A	2	3	0.3	0.4	4.4	4.8	N/A	6.7	N/A	5
T						68	6.8	N/A										
W	12/26/2001	200	74	35.4	N/A				5	1	0.3	0.6	4.6	5.2	N/A	6.7	N/A	5
T																		
F	12/28/2001					67	6.7	100								6.8	120	6
M	12/31/2001	596	796	32.5	244	64	6.7	240	1	7	0.3	0.7	5.4	6.1	75	6.5	140	5
T																		
W	1/2/2002	N/A	N/A	N/A	N/A	59	7.4	280	N/A	N/A	N/A	N/A	N/A	N/A	74	6.7	140	5
T						59	7.4	220								6.6	120	4.9
F	1/4/2002																	
M	1/7/2002	425	642	37.1	238	62	7.6	240	4	3	0.3	N/A	4.9	N/A	19	6.7	140	4
T																		
W	1/9/2002	156	118	34.2	202	64	6.9	-	11	3	0.3	0.8	17.6	18.4	123	6.3	60	9
T						65	6.9	140								6.5	120	7
F	1/11/2002					68	6.7	240								6.5	120	6
M	1/14/2002	368	460	35.7	238	68	6.7	260	5	4	0.3	0.7	121.6	122.3	48	7.1	80	7
T																		
W	1/16/2002	294	226	37.8	225										91			
T						60	7.5	240								6.7	200	8
F	1/18/2002					61	7.5	240								6.6	100	7
M	1/21/2002					59	7.1	240								6.5	140	10
T	1/22/2002	492	848	35.3	272	66	6.9	260	4	1	0.3	2.3	7.8	10.1	60	6.8	120	8
W	1/23/2002	344	428	30.7	223				4	7	8.6	11.3	1.5	12.8	125			
T																		
F	1/25/2002																	
M	1/28/2002	538	1040	25.8	269		6.8	140	4.4	3	0.3	0.9	10.8	11.7	48	6.6	100	8
T						67	7.3	200								6.5	100	7
W	1/30/2002	288	396	29.9	251	64	7.2	--	NA	1	0.3	1.7	12.1	13.8	47	6.7	120	7
T						63	6.9	240								6.5	100	8
F	2/1/2002																	
M	2/4/2002																	
T						58	6.8	120								6.8	100	8
W	2/6/2002	417	820	37.5	248	70	7	240	3	1	0.3	0.4	19.1	19.5	25	6.5	100	8
T						70	6.8	280								6.5	80	9
F	2/8/2002					61	7	140								6.6	100	9
M	2/11/2002	252	140	26.3	198				4	10	N/A	0.3	13.1	13.4	N/A			
T						61	7.3	180								6.6	100	9
W	2/13/2002	482	600	38.6	230	60	7	240	10	1	0.3	0.6	14.3	14.9	34	6.6	120	8
T																6.5	100	7
F	2/15/2002					60	7.3	260								6.6	120	6
M	2/18/2002																	
T	2/19/2002	617	796	31.6	207	70	6.7	240	5	1	0.3	0.8	13.7	14.5	33	6.7	100	9
W	2/20/2002	321	314	32.3	N/A				10	3	0.3	N/A	12.8	N/A	N/A			
T																		
F	2/22/2002																	
M	2/25/2002	662	1160	38	249				4	1	0.3	N/A	17.2	N/A	23			
T																		
W	2/27/2002																	
T																		
F	3/1/2002																	
average		371	519	33.8	235	64.4	7.04	212	5.4	3.2	0.88	1.6	16.6	20.6	58.9	6.71	125	5.64
stdev		165	326	3.64	22.8	3.73	0.27	49.8	3.1	2.6	2.07	3	28	31	34.7	0.20	35.3	3.15
min		99	74.0	25.8	198	58.0	6.60	100	1.00	1.00	0.30	0.30	0.60	4.80	19.0	6.30	60.0	0.10
median		356	518	34.5	238	65.0	7.00	240	4.2	3.00	0.30	0.7	12.1	13.4	48.0	6.70	120	6.30
max		662	1160	38.6	272	70.0	7.60	280	11.0	10	8.6	11.3	122	122	125	7.20	220	10.0

		CATION 3 (PROCESS TA			ANOXIC 1 (TANKS 1 & 2)			OXIC (TANKS 3,4,5,6 & 7)								ANOXIC 2		
				Field	TANK 1	LOCATION 4(Tank 2)		LOCATION 7 (TANK 3)		TANK 4	TANK 5	TANK 6	LOCATION 8 (TANK 7)			LOCATION 5 (TANK 8)		LOCAT
Day	Date	pH	MLSS	MLSS	DO	DO	NOx	DO	NH3	DO	DO	DO	DO	NH3	NOx	DO	NOx	DO
	(M/D/Y)		(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)	(mg/l)
F	11/9/2001																	
M	11/12/2001			7200	0.4	0.13		0.18		0.15	0.11	0.1	0.14			0.1		0.1
T	11/13/2001			12600	0.5	0.15		0.19		0.16	0.1	0.09	0.16			0.11		0.1
W	11/14/2001			N/A	0.51	0.17		0.2		0.15	0.112	0.1	0.15			0.12		0.13
T				2100	0.5	0.16		0.19		0.15	0.12	0.1	0.13			0.11		0.16
F	11/16/2001																	
M	11/19/2001			27000	0.55	0.19		0.19		0.2	0.15	0.11	0.2			0.1		0.13
T				7800	0.26	0.05		0.71		0.4	1.09	1.2	2.04			0.19		0.18
W	11/21/2001		N/A		0.21	0.47	0.7	0.77	11.3	0.49	0.96	1	1.4	23.3	0.6	0.25	0.3	0.2
T																		
F	11/23/2001			9000	0.25	0.43		1		1.1	0.96	1.1	1.6			0.2		0.17
M	11/26/2001		N/A	7800	0.16	0.1	0.5	1.02	21.2	1.13	1.15	1.49	2.22	24.1	0.4	0.19	0.3	0.19
T				N/A	0.21	0.15		1.14		1	1.12	1.43	2			0.15		0.24
W	11/28/2001		N/A	15000	0.18	0.13	0.3	1.16	8	1.1	1.15	1.5	1.97	16.5	0.4	0.2	0.3	0.2
T				15000	0.1	0.06		1.84		0.52	1.17	1.45	1.96			0.44		0.26
F	11/30/2001			3600	0.13	0.11		1.5		0.46	1	1.53	1.85			0.56		0.15
	12/1/2001																	
M	12/3/2001		N/A	3000	0.08	0.1	0.6	1	11.9	0.46	1.08	1.15	1.1	5.7	0.8	0.13	0.4	0.2
T																		
W	12/5/2001		N/A	3600	0.5	0.1	8.4	0.65	3.3	0.36	0.96	0.45	0.82	0.3	21.7	0.1	5	0.11
T	12/6/2001			4800	0.46	0.13		0.59		0.37	1	0.59	1.2			0.9		0.05
F	12/7/2001			4800	0.5	0.15		0.51		0.35	0.9	0.56	1			1.1		0.46
M	12/10/2001																	
T	12/11/2001			15000	0.46	0.1		0.67		0.45	1.2	0.65	1.25			1		0.62
W	12/12/2001			12000	0.54	0.09		1		1.1	1.2	1.01	0.9			0.35		0.65
T	12/13/2001		5900	N/A			3.1		N/A					N/A	9		9.6	
F	12/14/2001																	
M	12/17/2001		6640	6000	0.08	0.09	3.1	1	7.8	0.5	0.9	0.55	1.1	0.3	N/A	1	0.6	0.48
T				4800	0.28	0.28		3.42		1.1	2.73	3.51	4.08			0.95		0.1
W	12/19/2001		2060	12,000			0.7		5.7					0.3	N/A		0.6	
T																		
F	12/21/2001			9600	0.25	0.34		1.65		1.32	1.29	1	1.9			1.7		0.44
M	12/24/2001		8600	2400	0.4	0.38	0.6	1.4	17.6	1.3	1.29	1.01	1.79	10.2	N/A	1.76	0.6	0.42
T																		
W	12/26/2001		980	2700	0.4	0.38	0.6	1.45	37.1	1.36	1.09	1.01	1.95	10.3	N/A	1.75	0.6	0.4
T				6000														
F	12/28/2001			6000	0.25	0.36		1.6		1.31	1.15	1	1.92			1.01		0.48
M	12/31/2001		7580	6000			0.6		2.3					16.6	N/A		0.6	
T																		
W	1/2/2002		7440	6600	0.57	0.45	4.8	1.06	2	1.01	1.1	1.26	1.15	12.7	N/A	1.07	1	0.36
T				6000	0.55	0.49		1.03		1.41	1.03	1.46	1.41			1.11		0.27
F	1/4/2002																	
M	1/7/2002		N/A	6000	0.65	0.48	1.4	1.01	1.4	1.22	1.09	1.4	1.32	N/A	N/A	1.09	N/A	0.39
T																		
W	1/9/2002		5700	4200			10.4		0.8					0.3	N/A		N/A	
T				5400														
F	1/11/2002			3600	0.25	0.43		1.3		1.4	0.96	1.4	1.6			0.2		0.16
M	1/14/2002		4580	6000	0.58	0.33	0.7	0.92	0.6	1.05	1.02	1	1.05	13.7	N/A	0.75	N/A	0.68
T				6600														
W	1/16/2002		4080	6600			4.8		9.9					N/A	N/A		N/A	
T				--	0.57	0.35		0.96		1.04	1.16	1.27	1.32			1.02		0.48
F	1/18/2002			--														
M	1/21/2002				0.5	0.36		0.99		1	1.1	1.22	1.36			1.1		0.3
T	1/22/2002		6160	4800	0.45	0.4	0.6	0.88	0.6	1.12	1.11	1.26	1.32	16.1	N/A	1	N/A	0.3
W	1/23/2002		5520				0.7		N/A					2.5	N/A		N/A	
T															N/A			
F	1/25/2002																	
M	1/28/2002		6640	4800			1.5		0.4			1.04	1	N/A	N/A		N/A	
T				4920	0.2	0.18		1.06		1.07	1.03	1.14	1.1			1.1		0.33
W	1/30/2002		4820	--	0.22	0.19	0.6	0.51	7.1	1.06	1.12	1.31	1.24	0.3	N/A	0.4	N/A	0.37
T				4800	0.2	0.17		0.58		1.1	1.27	1.27	1.26			0.51		0.33
F	2/1/2002																	
M	2/4/2002																	
T				--	0.19	0.26		1		1.21	1.21	1.2	1.1			0.46		0.37
W	2/6/2002		190	4800	0.4	0.14		2.01		2.04	1.82	1.18	1.2			0.58		0.45
T				--	0.3	0.25	5.5	0.9	7.4	1.02	1.1	1.09	1.05	10.5	N/A	1.01	N/A	0.75
F	2/8/2002			5400	0.31	0.26		1.1		1.04	1.13	1.04	1.02			1.02		0.21
M	2/11/2002		5800				1.2		0.7					N/A	N/A		N/A	
T				6000	0.37	0.25		0.46		1.04	1.07	1.26	1.27			0.13		0.37
W	2/13/2002		9920	6000	0.34	0.24	1.2	0.41	6.6	1.02	1.1	1.29	1.2	6.6	N/A	0.2	N/A	0.36
T					0.34	0.22		0.47		1.05	1.02	1.2	1.26			0.2		0.32
F	2/15/2002			6000	0.39	0.25		0.45		1.12	1.06	1.19	1.2			0.21		0.3
M	2/18/2002																	
T	2/19/2002		7520	--	0.89	0.9	0.6	1.01	0.3	1.02	1.08	1.15	1.12	17.6	N/A	0.3	N/A	0.3
W	2/20/2002		7740				0.6		15.1					5.6	N/A		N/A	
T																		
F	2/22/2002																	
M	2/25/2002		10760				4.3		0.5					15.4	N/A		1.1	
T																		
W	2/27/2002																	
T																		
F	3/1/2002																	
average			5932	7077	0.37	0.25	2.23	0.96	7.48	0.89	1.03	1.07	1.31	9.95	5.48	0.62	1.62	0.31
stdev			2689	4528	0.17	0.16	2.65	0.58	8.64	0.43	0.43	0.55	0.67	7.69	8.63	0.49	2.70	0.17
min			190	2100	0.08	0.05	0.30	0.18	0.30	0.15	0.10	0.09	0.13	0.30	0.40	0.10	0.30	0.05
median			6030	6000	0.37	0.22	0.70	1.00	6.15	1.04	1.09	1.15	1.25	10.30	0.70	0.46	0.60	0.30
max			10760	27000	0.89	0.90	10.4	3.42	37.1	2.04	2.73	3.51	4.08	24.1	21.7	1.76	9.60	0.75

		(TANKS 8, 9 & 10)					COMMENTS
		ION 9 (TANK 9)		LOCATION 6 (TANK 10)			
Day	Date	NOx		DO	NOx		
	(M/D/Y)	(mg/l)		(mg/l)	(mg/l)		
F	11/9/2001						
M	11/12/2001			0.09			
T	11/13/2001			0.08			Blower tripped and was reset
W	11/14/2001			0.09			
T				0.11			
F	11/16/2001						
M	11/19/2001			0.08			Wasted 100 gallons of Sludge
T				0.16			
W	11/21/2001	0.30		0.19	0.30		
T							
F	11/23/2001			0.15			Permeate Pump Tripped, Process Blower Tripped - Both rese
M	11/26/2001	0.60		0.16	4.50		
T				0.15			High Vacuum Alarm; Performed Cleaning
W	11/28/2001	0.4		0.15	0.4		Performed Clean; Wasteed 200 gallons of Sludge
T				0.17			
F	11/30/2001			0.13			
	12/1/2001						Performed Cleaning
M	12/3/2001	0.60		0.5	2.10		High Vacuum Alarm, unable to reset
T							
W	12/5/2001	14.90		0.2	12.40		High Vacuum Alarm
T	12/6/2001			0.11			High Vacuum Alarm; No Air to Zeeweed Tank
F	12/7/2001			0.5			High Vacuum Alarm
M	12/10/2001						
T	12/11/2001			0.5			Unit out of Alarm: Wasted 200 gallons of Sludge
W	12/12/2001			0.47			Performed Cleaning: Wasted 200 gallons of Sludge
T	12/13/2001	8.00			8.20		Wasted 100 gallons of Sludge
F	12/14/2001						
M	12/17/2001	0.6		0.34	0.6		
T				0.1			
W	12/19/2001	0.70			0.7		
T							
F	12/21/2001			0.36			
M	12/24/2001	0.60		0.38	0.6		
T							
W	12/26/2001	0.70		0.38	0.6		
T							
F	12/28/2001			0.4			
M	12/31/2001	0.60			0.6		
T							
W	1/2/2002	0.60		0.44	0.6		
T				0.36			
F	1/4/2002						
M	1/7/2002	N/A		0.33	0.7		High Vacuum Alarm
T							
W	1/9/2002	N/A			15.3		High Vacuum Alarm 10+ psi; #5 and #9 sample ports cloggec
T							
F	1/11/2002			0.11			
M	1/14/2002	N/A		0.65	0.6		
T							
W	1/16/2002	N/A			1.2		
T				0.35			
F	1/18/2002						
M	1/21/2002			0.26			
T	1/22/2002	N/A		0.28	0.5		
W	1/23/2002	N/A			2.7		
T							
F	1/25/2002						Process Blower Repaired
							Maintenance Clear
M	1/28/2002	1.20		0.24	1.2		
T				0.27			
W	1/30/2002	N/A		0.22	0.6		
T				0.21			
F	2/1/2002						Maintenance Clear
M	2/4/2002						
T				0.21			
W	2/6/2002			0.3			
T		N/A		0.2	5		
F	2/8/2002			0.21			Maintenance Clear
M	2/11/2002	N/A			0.6		
T				0.2			Maintenance Clear
W	2/13/2002	N/A		0.17	3.6		Maintenance Clear
T				0.15			
F	2/15/2002			0.17			Tank 5 & Tank 9 aeration for 30 sec. Every 5 minutes
M	2/18/2002						
T	2/19/2002	N/A		0.17	0.6		
W	2/20/2002	N/A			1		
T							
F	2/22/2002						
M	2/25/2002	N/A			0.8		
T							
W	2/27/2002						
T							
F	3/1/2002						
average		2.29		0.25	2.54		
stdev		4.31		0.14	3.83		
min		0.30		0.08	0.30		
median		0.60		0.21	0.70		
max		14.9		0.65	15.3		

APPENDIX B

March 1, 2001 — Set up and commissioning work started

April 10, 2001 — Set up complete; pilot started in modified batch mode to reach target of 8,000 mg/L

April 10 to May 8, 2001 — Initial start up, system seeding and acclimation

May 8, 2001 — Concentration of 8,000 mg/L had been obtained in membrane tank

May 9 to May 25, 2001 — Phase 1 – Direct filtration

May 9, 2001 — Process set points:
Flux = 11 gfd
Sludge wasting = none
Relax frequency = 10 minutes
Relax duration = 30 seconds
Maintenance cleanings = 1 (NaClO)
Chemical dose = none
Chloramines in backpulse = none
Air = 16 cfm, cyclic with on/off intervals of 10 sec
Recirculation rates = 15/15 gpm for inner/outer

Note: MLSS results inconsistent: analytical sampling not completed

May 26 to July 25, 2001 — Phase 2 – Increased recirculation rate

Increased outer recirculation rate from 15 gpm to 25 gpm
Flux = 11 gfd
Relax frequency = 10 minutes
Relax duration = 30 seconds
Maintenance cleanings = 1 (NaClO)
Air = 16 cfm, cyclic with on/off intervals of 10 sec
Recirculation rates = 15/25 gpm for inner/outer

Feed pump and line broke – line repaired, pump replaced
MLSS results inconsistent
No sludge wasting
System shut downs and power failures
Low air to membranes – supplemental blower sent to site
Clogging between tanks 2 & 3 – fixed by operator

July 22, 2001 — ZENON rep on site for 3 days
Installed blower
1,500 gallons sludge wasted
blower vanes and air filters replaced

July 25, 2001 System returned to service with increased air flow to membrane tank (now at 30 cfm)

July 25 to August 27, 2001 Phase 3 – Increased air to membranes

Flux = 11 gfd
Relax frequency = 10 minutes
Relax duration = 30 seconds
Maintenance cleanings = 1 (NaClO)
Air = 30 cfm, cyclic with on/off intervals of 10 sec
Recirculation rates = 15/25 gpm for inner/outer

August 27, 2001 ZENON representative on site
Aeration flow to tank #8 was shut off, creating a larger anoxic and smaller aerobic zone in the overall tank scheme. (Tanks 1 and 2 remain anoxic, tanks 3 to 7 remain oxic, tanks 8, 9 and 10 now anoxic).

Rerouting membrane tank overflow from tank #1 to tank #3 (anoxic to aerobic) Note: on November 1, this change was fully made

August 28, 2001 DO readings taken by ZENON representative on August 28th showed the following results:
Tank #1 (anoxic) 0.20 mg/L
Tank #2 (anoxic – end of first zone) 0.17 mg/L
Tank #7 (oxic – end of aerobic zone) 2.20 mg/L
Tank #8 (anoxic – start of second anoxic zone and feed supply to inner recirculation loop) 1.80 mg/L
Tank #10 (anoxic – end of second anoxic zone and feed/outer flow loop supply to ZeeWeed® membrane tank) 0.50 mg/L

August 28 to November 6, 2001 Phase 4 – Change in tank configuration

September 26, 2001 Vacuum increased to 2 psi
Vacuum continued to climb to over the next 9 days to 4.4 psi

October 4, 2001 First few weeks in October, vacuum remained high – operators performed daily maintenance cleans with NaClO to reduce vacuum

mid-October, 2001 Representative from O'Brien and Gere on-site
OBG rep installed air valves into anoxic tanks to help with mixing

October 23, 2001 Target Conditions:
Feed flow = 5 gpm
Permeate flow = 5 gpm

Recirc Pump #1 (inner) = 15 gpm
Recirc Pump #2 (outer) = 25 gpm
Overflow at 20 gpm (dif b/w pump 2 and feed)
25 cfm air
air cycling at 10/10
10 sec/30 min relax cycle
maintenance cleans 3 x /week with NaOCl
6 cfm to aerated tanks
10 g/L MLSS by wasting

System on high vacuum

Mechanical problems:

not enough air to membranes
25 scfm required for membranes
can get 17 scfm from current blower
new blower sent – not working?
Veins reversed, factory defect, blower not installed, isn't working
lack of mixing in anoxic tanks
submersible pumps not working
sent equipment to pulse air
influent bag filter housing – may have taken mesh out.
How long running like this?

Information from OBG representative

Problem with system
high vacuum alarm
for one week on alarm
maintenance clean every day – sodium hypochlorite
blower situation
sampling ports clogged

Information from site personnel

High vacuum
Cleanings NaOCl – maintenance clean – add 1 qt to CIP tank
Membrane aeration 17 scfm
Relax O.K.
MLSS 10,200 mg/L
Wasting approx 100 gal /day

Aerate system for few hours or overnight

October 24, 2001 Still getting alarms
Timer was installed this morning for anoxic tank

October 25, 2001 low level and high vacuum alarms

ZW-tank aerated overnight
Feed pump working
15,000 mg/L MLSS

Information from site peronnel
Strainer was cleaned – not much around
Membrane tank is aerating – confirmed by Bill

October 26, 2001 high vacuum alarm
Valve 4 closed – open again now

October 29, 2001 high vacuum alarm

October 31, 2001 Site visit by ZENON rep to determine cause of high vacuum

Aerating the membrane overnight
Vacuum dropped to 1.5 psi from 10 psi

November 1, 2001 Aerobic tanks aerated at 2 cfm instead of 6 cfm
Air to membranes at 10 – 15 cfm instead of 25 cfm
Pump skid #1 at 1 gpm instead of 15 gpm
Basket strainer plugged very badly
High vacuum alarm – after aerating, vacuum at 3" Hg
Feed pump not in center of tank

Small blower on system – giving 10 cfm
8 – noon, 3" Hg – 15 " Hg
new blower veins and filters being sent
aeration in anoxic tanks installed
not running sprayer pump – is this okay?

November 2, 2001 sent today – veins for blowers
filters
fittings to connect air to other blower
blower for ZW-10 – does it give 5 – 10 cfm?
palette in pilot shop

mixing for anoxic tanks – check timing
will start testing next week when system operating
aerobic tanks not always at 6 cfm
any procedures that may be required should be left with them
train – maintenance clean and daily checks

November 5, 2001 No sprayer nozzle on ZW tank –send
Lots of foaming
Blower working

Check valve for blower
Running at 22 cfm

November 6, 2001 Site is pretty messy
Tank #1 – no aerators
Most of flow still going to tank #1 from ZW tank
(tried to change on Aug 27 when ZENON rep was on-site to feed tank #3 from ZW tank)
mixing of tank 1 poor
may be able to change feed location

second blower was installed to increase the air flow up to the requirement of 25 scfm

November 7 to February 27, 2002 Phase 5 – Change in tank configuration II

November 7, 2001 Running fine
Air 25 cfm to membrane tank
Logsheets submitted

November 8, 2001 Unit has not operated for more than 24 hours at a time
overflowing, foaming over
leak in camlock
No mixing in first anoxic tank
put pump in for mixing
make sure we have back pressure on it
everything below liquid level
mixing pump for anoxic tank
overflow for tank 3

correct overflow from ZW-tank to tank 3
Recirc pump in tank 1 installed to mix contents
Running at 3 gpm, 1" Hg, air at 15 cfm, recirc at 12 gpm to ZW-tank, 15 gpm to tank 1
Foaming a little bit

November 13, 2001 blower in aerobic tanks down last night – reset
System off when operators in, in morning
System at 4"Hg
Wasted 100 gal
Power failure last night
Low level alarm
Ammonia conc up to 5 mg/L
may be due to loss of air to the anaerobic tanks caused by a power failure

November 14, 2001	conference call with Bhavani, Lowell and Sami Operational - changes on site mixing in tank 1 recirc line from ZW tank to tank 3 (not done properly before air flows in aerobic tanks Analytical Higher ammonia due to blower down Monday night D.O. probably not correct – operators not taking samples correctly – from sample valves not top of tank Only need D.O. to ensure process correct Bhavani to go to site tomorrow to measure D.O. -if D.O. is O.K., cut back # of D.O. samples taken - take D.O. samples of last tanks (2, 7, 10) in trains (3 samples) D.O. meters on site – Cory to tell Bhavani which meter Greg used Sami suggesting getting a standard D.O. to calibrate D.O. meters – or use Winkler method
November 16, 2001	everything going well operationally
November 20, 2001	Spoke to Bill Doubleday Reading from 11/19/01 Alarm last Thursday (11/15/01) due to power glitch Problem with level transducer – Greg troubleshooting with Bhavani Bhavani turned recirc to 17 gpm – (later phone conversation with Bill Doubleday indicated that he increased it to 20 gpm)
November 21, 2001	Conference call with Bhavani Samples taken to lab Wasted 150 gal (MLSS at 15 000 mg/L) Conference call with Steve W., Bhavani Rathi, Lowell, Cory: Dissolved Oxygen #'s did not seen correct form operators for D.O. diaphragm valves not working well – air flow to tanks decreasing need to do daily check of numbers – get from Cory recirc reset 25 gpm + 17 gpm blower at 25 cfm on membrane skid level controller was working operators have not taken samples to lab Operational data Operational data from operators – Bhavani will ask D.O.'s every day until we get consistent data

	7800 mg/L nitrates 5 mg/L NH_3 0.2 mg/L Nitrate conc slightly high – may be because of low recirculation rate between tanks 8 and 1
November 23, 2001	Permeate pump tripped out – reset at 2:00 – caused by main plant generator overload Process blower tripped out – reset
November 26, 2001	all fine (MLSS low)
November 27, 2001	instructed plant personnel to increase recirc from Tank 8 to Tank 1 to 20 gpm informed Bill already did this 11/20/01 high vacuum alarm last night maintenance clean today with chlorine informed that plant personnel had not conducted maintenance clean since Greg left asked plant personnel to conduct maintenance cleans 3 x /week MLSS 15 000 mg/L according to Bill – wasted 150 – 200 gal (Eric got MLSS reading of 7800 mg/L) system off on a high vacuum alarm aerated for several hours vacuum decreased to 1.5" Hg maintenance cleans were reinstated conducting maintenance cleans three times a week, twice with chlorine and once with citric acid
November 29, 2001	Nitrates up a little Sami thinks we should be getting TN < 3 or 4 mg/L Wait until we get lab results before changing conditions
December 3, 2001	System shut off due to a high vacuum alarm High rate of membrane fouling due to lack of aeration to the membranes
December 4, 2001	High pressure alarm – started yesterday Been doing maintenance cleans (not enough air to membranes) high ammonia approx 5 mg/L nitrate 2 mg/L tanks a little low on air

	one blower for supplemental two blowers for membrane
December 5, 2001	Blower to membranes check bypass on both blowers (?) leaks between blower and rotameter (?) air is cycling through muffler getting some air in membrane tank kink in hose from rotameter
December 6, 2001	Blower problems Aerator flush – how to do it? Pump in permeate break tank – discharge of the pump – connect to the 1" line air right after air rotameter With large blower discharge T may be 1" Connect to air line May blow out obstruction with 30 cfm Or turn off 9 ball valves, take 10th to air line Greg – could take apart air line after rotameter and see if air comes through (+ measure)
December 11, 2001	Bill – tried to do tasks on fax Got air into membrane Last week? Not running 3 parts ammonia 7 parts nitrates blower problems resolved
December 12, 2001	Pilot made it through the night No samples from lab on permeate
December 13, 2001	Pilot still running Nitrates 6 mg/L NH_3 0.1 Recirc at 20 gpm? Confirm
December 19, 2001	Call from Bhavani – everything went well on site recirculation rates were adjusted aeration to Tank #4 was low.
December 21, 2001	Call with Bhavani Tank #1 sludge blanket, likely because no air to tank mixing from pump have operator lift up pump to check for mixing

	recirc rates – adjust vacuum readings – log sheets
January 5, 2002	Pilot off on high vacuum alarm Vacuum back down to 3 psi after aerating overnight and a maintenance clean
January 7, 2002	System off on alarm over weekend
January 9, 2002	Bhavani – no new information from lab
January 10, 2002	Bhavani – talked to Bill Doubleday – plant running
January 23, 2002	Tank #9 sample port plugged Vacuum not checked Recirc rates not checked Air looks good Aerobic tanks – no air going through valves membrane tank @ 25 cfm rotameters on each tank 2 way valve pneumatic – anoxic – if open, may reduce air to aerobic tank blower on? Air rotameter on discharge of blower Vanes need to be replaced? Pump spinning Large 1 ½” ss valve on discharge may be closed
January 23, 2002	OBG representative on site sampling ports to Tanks 5 and 9 were plugged blower supplying air to the aerobic tanks was not working properly. Aeration to Tanks 5 and 9 increased to 30 seconds every 5 minutes to break up the sludge blankets in these tanks and clear the sampling ports.
January 25, 2002	Pilot system off of high vacuum alarm Vacuum back down to 3 psi after aerating overnight and a maintenance clean System operated between 2 and 3 psi until the end of January New veins for the blower were sent to site and installed
January 29, 2002	One elevated nitrate level was noted (7 mg/L) MLSS concentration was low

January 30, 2002 conditions for the system were confirmed at:
6 cfm air to the aerobic tanks
recirculation rates of 20 and 25 gpm
MLSS concentration of 4,800 mg/L

February 6, 2002 Pilot went off on high vacuum alarm
Vacuum returned to 3.5 psi after aerating overnight and conducting a maintenance clean
system continued to run, without alarms, until late February

MLSS concentration low, 4 800 mg/L

February 13, 2002 MLSS concentration increased to 6 000 mg/L and remained there until February 20

February 26 to 28, 2002 On-site visit
Unit not operating
Sludge blankets in tanks 5 and 9 not broken up
Anoxic tanks aerated continuously with 6 cfm of air per tank overnight to break up the sludge blankets
System restarted and the vacuum close to 15” Hg
System off on high vacuum alarm

Problems:
compressed air supply was not set at 80 psi,
the permeate turbidimeter was not working properly
the recycle pumps were not running
solenoid valves needed to be changed
the level logic was incorrect
the permeate pump was pulling a lot of air
a pneumatic valve on the permeate line was leaking,
the chlorine injection into the backpulse tank was not working
the membrane vacuum was high

Maintenance clean conducted with 500 mg/L of chlorine (backpulsing and relaxing the membrane for 60 and 300 seconds respectively x 10)
Soaked overnight in chlorine.

February 27, 2002 Vacuum still high
Recovery clean with 2000 mg/L of chlorine started
pneumatic valve was changed
chlorine injection pump was replaced
the recycle pumps were reset and started working
the compressed air supply was increased
the level logic was reset

	the solenoid valve that controlled the cyclic aeration to the membranes was replaced. Soaked membranes in chlorine overnight system was still going off on high vacuum alarm at a flow rate of 4 gpm backpulse pressure had decreased to 3 psi from 6 psi.
February 28, 2002	Agreement with ZENON and O'Brian and Gere that system should be left in clean water and soaked in citric acid once this product has been delivered on site Site personnel report nitrate levels between 5 and 10 mg/L for the month of February (these were higher than previously seen in study) In February, all MLSS concentrations recorded were lower than target
March 1, 2002	Recovery clean with 10 g/L citric acid started
March 5, 2002	System started with low vacuum
March 6, 2002	System ran for a few hours with vacuum less than 1" Hg Problems with the feed flow System off
March 8, 2002	Troubleshooting Thornton controller and feed flow
March 13, 2002	Reprogrammed Thornton controller
March 27, 2002	Decision made to shut down pilot

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New York State Energy Research
and Development Authority
17 Columbia Circle
Albany, New York 12203-6399

toll free: 1 (866) NYSERDA
local: (518) 862-1090
fax: (518) 862-1091

info@nyserda.org
www.nyserda.org

IMMERSED MEMBRANE BIOREACTOR PERFORMANCE EVALUATION
TWELVE PINES SEWAGE TREATMENT PLANT; SUFFOLK COUNTY, NEW YORK

FINAL REPORT 04-04

STATE OF NEW YORK
GEORGE E. PATAKI, GOVERNOR

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