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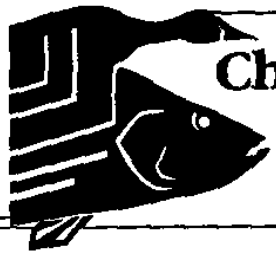
Biological Nitrogen and Phosphorus Removal in Oxidation Ditch and High Nitrate Recycle Systems



Chesapeake Bay Program

Biological Nitrogen and Phosphorus Removal in Oxidation Ditch and High Nitrate Recycle Systems

by

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EXECUTIVE SUMMARY

Full scale process modifications were conducted in an oxidation ditch system to understand the kinetics of biological nitrogen and phosphorus removal in activated sludge systems which may use high nitrate or internal recycles. Design parameters and operating strategies with multiple control parameters were evaluated over a two-year period to help increase the efficiency of year round biological nutrient removal in temperate climates. The full scale evaluation has helped identify the extent of nitrogen and phosphorus removal which may be achieved, and the corresponding changes in the kinetics of reactions when nitrate recycle rates are increased to enhance denitrification in activated sludge systems.

The Bowie WWTP, located on the Patuxent River in Prince George's County, Maryland, was modified by the Research Division of Virginia Polytechnic Institute and State University to maximize biological nitrogen and phosphorus removal and satisfy year round permits of 1 mg/L for phosphorus and 6 mg/L for total nitrogen (monthly averages) without the use of chemicals. The research was sponsored by the Maryland Department of the Environment and the USEPA Chesapeake Bay Office.

The Bowie WWTP is an oxidation ditch system which receives 2.2 MGD of flow. The facility was modified in Phases to investigate biological nitrogen and phosphorus removal in oxidation ditches and optimize process design and operations within the constraints of the structures available. The final design utilizes two oxidation ditches connected in series with a side stream anaerobic cell. The anaerobic cell receives only 40 to 60 percent of the raw influent flow. The mixed liquor recycle is set at 30 to 45 percent. Therefore, the VT2-BNR process, as developed for this facility, combines the capacity for biological nutrient removal in oxidation ditches (i.e., without a separate

anaerobic cell) and enhances phosphorus removal capacity with the anaerobic cell. A schematic of the plant is presented in Figure 1.

Two oxidation ditches were necessary to maintain the MCRTs essential for winter nitrification without overloading the clarifiers. With only one oxidation ditch in operation, the ammonia-N levels averaged in excess of 5 mg/L in winter (1988-89). When two ditches were put in operation (in series or in parallel), effluent TKN averaged 1 mg/L and total nitrogen averaged 3 mg/L.

By controlling the aerobic, anoxic and anaerobic volumes inside the oxidation ditches, it was possible to enhance biological phosphorus removal and average less than 2 mg/L in the effluent with about 4 percent phosphorus in the biomass on a volatile solids basis. To average less than 1 mg/L, it was necessary to utilize an anaerobic cell whose volume was 10 percent of the combined volume of the oxidation ditches. Effluent phosphorus levels averaged 0.5 mg/L in warm weather and 0.6 mg/L in cold weather. Phosphorus removal has continued to improve with improvements in process control. The system is expected to average less than 0.5 mg/L annually at MCRTs between 15 and 35 days. Because of a phosphate detergent ban, the influent BOD_5/P ratio averages 30/1 and the COD/P ratio averages 70/1.

The plant utilizes gravity thickening and belt filter presses for solids dewatering. Less than 10 percent of the waste sludge phosphorus is recycled from the gravity thickeners. Precautions are taken to prevent excessive solids losses in the supernatant. Non-potable water is not added to gravity thickeners to minimize the volume of supernatant.

The phosphorus content of the plant recycle increases the influent phosphorus concentrations by 25 percent. This is because the belt filter presses are overloaded and, therefore, are not very efficient in dewatering the solids. The BOD_5/P ratio in the influent to the oxidation ditch after it is mixed with the plant recycle averages 22/1.

The research has identified several parameters which can enhance process control in oxidation ditches. These include parameters for the control of filamentous bacteria and alternatives to dissolved oxygen measurement to

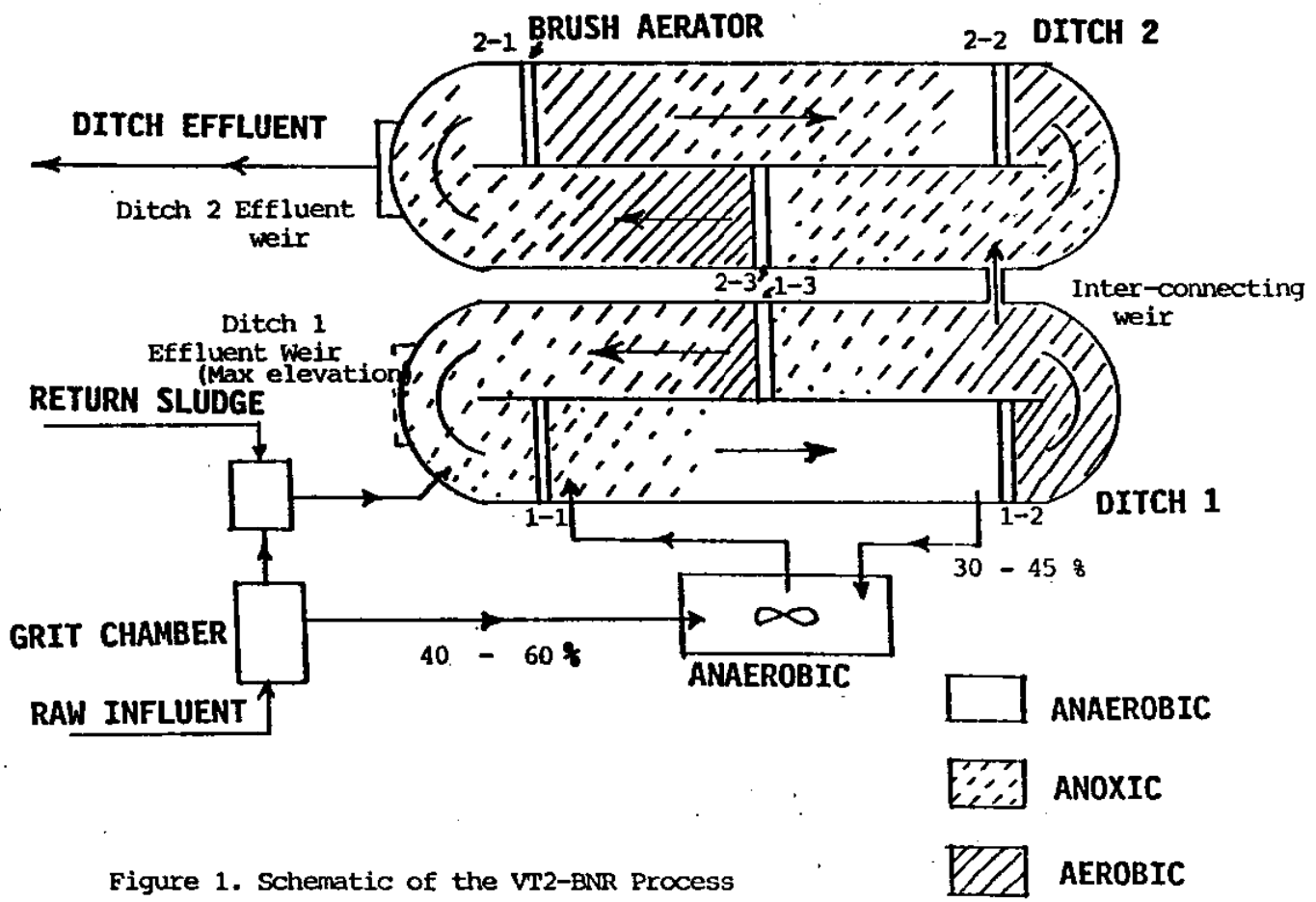

Figure 1. Schematic of the VT2-BNR Process

control aeration. SVI and filamentous bacteria can be kept in check when strict control is maintained over the anaerobic, anoxic and aerobic volume fractions inside the oxidation ditch. Since dissolved oxygen levels measure less than 0.5 mg/L at all accessible points in the ditch, alternative process control parameters were developed to correlate the pounds of oxygen injected to the process performance. These include secondary and aeration effluent alkalinity and turbidity measurements.

The available influent alkalinity at municipal wastewater treatment plants (influent alkalinity + alkalinity released during deamination of organic nitrogen) is a function of the influent TKN concentration. The effluent alkalinity can be related to the influent alkalinity as follows:

Effluent Alkalinity = Available Influent Alkalinity - 3.57 (Influent TKN)
+ Effluent N forms effect

Effluent N forms effect = 3.57 (Eff Ammonia-N) - 3.57 (Eff Oxidized-N).

If non-biodegradable forms of organic nitrogen are present in the influent, they will be removed in the effluent or in the waste sludge. The contribution of non-biodegradable organic nitrogen forms to the first and second terms on the right hand side of the equation are equal. Therefore, they cancel each other out. The TKN test captures a very high percent of the unoxidized organic and inorganic nitrogen forms. However, if a significant portion influent unoxidized nitrogen is of a form that it is biodegradable but is not captured in the TKN test, the equation will have to be expressed in terms of unoxidized nitrogen forms instead of TKN.

Nitrification and denitrification were controlled by adjusting the volume of biomass present under aerobic and anoxic conditions. An operating range was set up to maintain the effluent alkalinity between 72 and 75 mg/L to maintain excellent nitrogen removal. Aeration was controlled by changing the depth of submergence of the brush aerators (altering the liquid level in the oxidation ditches) or by adjusting the length of operation of the brush aerators with timers.

The effluent turbidity increased with alkalinity when the amount of air injected was not adequate for phosphorus removal. In oxidation ditches, which typically operate with very high internal recycle flows, biological phosphorus and nitrogen removal can take place simultaneously. Therefore, a short term reduction in the oxygen supply will affect both phosphorus and nitrogen removal. Effluent ortho phosphorus will also increase when dissolved oxygen and nitrate levels increase in the anaerobic sections of the oxidation ditch or in the anaerobic cell. The dissolved oxygen and nitrates were kept in check with alkalinity and turbidity measurements, and with occasional dissolved oxygen and ORP measurements in the anaerobic cell. Nitrogen and phosphorus removals were maximized by minimizing the volume of biomass maintained under aerobic conditions such that it was barely sufficient for nitrification and BOD removal. This helped maximize the anoxic and anaerobic volumes for nitrogen and phosphorus removal. The mixed liquor suspended solids had to be increased in winter to maintain adequate aerobic, anoxic and anaerobic MCRTs within the volume available.

The net annual saving after implementing the biological nutrient removal process was calculated as $57,000 at a flow of 2.2 MGD. This was observed in spite of an increase in nitrification of 10 mg/L. Of this, $30,000 was related to the savings in ferrous sulfate and was reflected directly in the operating budget. Other savings were calculated indirectly. The cost for adding supplemental alkalinity would have averaged $37,500 (as 30 percent caustic soda) if chemical phosphorus removal was practiced. The sum total of costs for aeration, recycle pumping and aerobic digestion decreased by $7,500. Sludge production after solids dewatering remained at the same level for a number of reasons including the fact that aerobic digestion was practiced prior to biological nutrient removal when the plant was operated at a lower MCRT. An additional $5,000 of savings in supplemental alkalinity was observed because of excellent denitrification which reduced effluent nitrate concentrations from 6 mg/L to 2 mg/L. Laboratory technician time and equipment expenses are expected to increase by $23,000 each year because of additional tests which have to be performed for the new NPDES permits for nitrogen removal.

ACKNOWLEDGEMENTS

On behalf of all the authors, I would like to thank the staff at the Occoquan Lab for their assistance with the data collection and analysis. Their help during those stages when little was known about the response of the biological system to process changes was invaluable. This study has helped us gain a better understanding of the response of biological systems to unsteady state conditions created by changes in temperature, flows and mechanical failures. My gratitude towards Dave Sirois, Harry Post and Barb Angelotti for their assistance in completing the project in spite of a deficit in funds provided for the laboratory analysis cannot be expressed in words. I would like to thank Rick Osgood for his untiring effort with on-site analysis and a level of dedication scarcely found elsewhere.

It would not have been possible to complete this study, had the staff at the Bowie WWTP not taken it up on themselves, the task of conducting the day to day operations. Special thanks need to be mentioned to Joe Schneider and Bill Kreitzer for supervising operations at the plant, to Nick Grogard for assisting with the laboratory work and process operations, to Josh and Sam Clark for paying particular attention to the vagaries of weekend flows, and to Phil, Bryan and other operators for chipping in when more pieces of equipment appeared to be failing rather than working.

We would also like to thank Jim Welch, Elmer Brown and Chuck Edwards at the Maryland City Water Reclamation Facility. Process modifications at the high rate Maryland City WRF (4 hour HRT) provided information which was invaluable in extending our understanding of activated sludge processes. The contrast between the high rate process at Maryland City and the low rate process at Bowie helped us refine operating concepts and control strategies to cover a wide range of activated sludge processes.

The evaluation of oxidation ditch systems would not have been as comprehensive, had we not had the opportunity to work with other oxidation ditch systems at the Patuxent WRF and Broadneck Facilities. Dan Rumke and Bill Shreve provided us with excellent opportunities to evaluate full scale situations, and helped us gain an insight into the instrumentation and control features which should accompany any nutrient removal upgrade.

I would also like to thank Buchart-Horn and Anne Arundel County for providing the opportunity to share experiences in nutrient removal in a variety of low rate systems (single point aeration oxidation ditch and Schreiber Counter Current Diffuser systems) and for pioneering work on automation and control in nutrient removal systems. If the proof of the pudding is in eating it, the Patuxent plant has consistently achieved effluent total nitrogen levels of 1 mg/L with single stage anoxic zone, and without the use of methanol or activated carbon treatment because of the versatility of the instrumentation. It is rare that a combination of operations personnel, consulting engineers and university researchers can work together to fine tune a biological system to whatever level of performance they may desire. It is our sincere wish that the staff at Patuxent, Maryland City, Broadneck and Bowie facilities publish some of the data which will be generated in future. The research reported here can at best be described as the introduction to the continuing effort at these facilities.

The initiative for biological nutrient removal modifications would not be present, had the State of Maryland not pressed for the research and the modifications. The University will be indebted to Secretary M. Walsh, Mr. J. Rein, Mr. K. Patel and Mr. M. Jiang for their support. The "additions and subtractions" to processes to create the right habitat for the bacteria would not have been possible without the administrative support from Tony Allred, Patsy Allen and from the City of Bowie. Last but not least, we would like to thank Joe Macknis and others at the Chesapeake Bay Office of the EPA for the opportunity to conduct full scale modifications and research at the Bowie WWTP.

On a more personal side, I would like to thank Andy Randall and Ken Brannan for "breaking the ice" with the staff at the plants to the extent that the operators went along with even the most absurd changes to operations. Jake Bair and Lenny Gold gave me the opportunity to develop training programs through the Maryland Center for Environmental Training to help make a smooth transition for all plants adopting biological nutrient removal in future. I will always be indebted to Dr. Randall for his initiative in selecting an oxidation ditch system to break new ground in research and process engineering, and to Dr. Grizzard for keeping the project together. Had it not been for Tom's (Dr. Grizzard's) effort, the crucial phase of this research - Phase V - would never have materialized.

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INTRODUCTION

The Bowie WWTP is located on the Patuxent River in Prince George's County, MD. The plant was designed in 1982 to operate as an oxidation ditch system and satisfy a BOD_5 (5 day Biochemical Oxygen Demand) and TSS (Total Suspended Solids) permit of 30 mg/L each. In 1985, facilities were installed for chemical phosphorus removal and caustic soda addition and the plant was issued a total phosphorus permit of 1 mg/L.

As part of the Chesapeake Bay Agreement to which the State of Maryland is a signatory, several plants in Maryland will be required to reduce the nitrogen and phosphorus levels in their effluent. To examine the feasibility of biological nutrient removal (biological nitrogen and phosphorus removal), the Bowie WWTP was selected because there were few full scale studies on biological nitrogen and phosphorus removal in oxidation ditches under temperate weather conditions.

The Bowie WWTP was upgraded for biological nutrient removal (BNR) in several phases. For purposes of the final report, these phases have been designated as follows:

Phase I Operation of the facility with one oxidation ditch and chemical P removal (1985 to Aug 1987);

Phase II Analysis of aeration capacities for nitrogen removal while continuing with chemical P removal (Sep 1987 to Jul 1988);

Phase III Operation of the facility with one oxidation ditch after discontinuing chemical P removal (Sep 1988 to Jan 1989);

Phase IV Operation of the facility with two oxidation ditches in parallel and no chemical P removal (Feb 1989 to May 1989);

Phase V Operation of the facility with two oxidation ditches in series - VT2 BNR process (Jun 1989 to end of project, Jun 1990); and

Phase VI Permanent upgrade to VT2 BNR process (Planned for FY 1990-91).

During Phase II, the amount of air supplied to the oxidation ditch system was not adequate for nitrification from September 1987 to May 1988. The effluent ammonia-N averaged above 10 mg/L. In June 1988, the oxygen supply was increased to achieve good nitrification while maintaining a high level of denitrification.

During Phase III, chemical P removal was discontinued to examine the potential for removing phosphorus biologically. The phosphorus levels in the effluent averaged 2 mg/L. Problems with process control made it difficult to maintain an adequate and consistent anaerobic volume in the oxidation ditch. This prevented further improvements in effluent phosphorus.

The plant was able to maintain total nitrogen levels below 5 mg/L when air and temperature did not limit nitrification. In November 1988, the MCRT was increased to maintain nitrification during cold weather. However, it was not possible to maintain complete nitrification and good denitrification in December 1988 and January 1989 because of a combination of factors:

i) the secondary clarifiers were overloaded at total MCRTs of 15 days;
ii) it was difficult to maintain adequate aerobic MCRTs for complete nitrification because of
   a) lack of sufficient aeration capacity to satisfy the demand during high flows;
   b) loss of anoxic MCRT when aeration was maximized during low flows (lack of adequate process control).

This resulted in a simultaneous increase in ammonia and nitrates in the effluent. The total nitrogen levels increased above 10 mg/L when liquid temperatures dropped below 18 Celsius.

During Phases I, II and III, the plant utilized the second oxidation ditch as an aerobic digester. Waste sludge was pumped to the aerobic digester when the solids dewatering system could not handle the amount of biomass generated. Between 5 and 15 percent of the biomass was pumped to the digester.

During Phase IV, both oxidation ditches were put in service in parallel to ensure excellent nitrogen removal in cold weather. The operating MCRT was increased well above that required for nitrogen removal to reduce the load on the solids dewatering system. Total Kjeldahl Nitrogen (TKN) levels averaged less than 1.5 mg/L and Total Nitrogen (TN) levels averaged less than 3 mg/L. Improvements in process control increased biological P removal. The effluent Total Phosphorus levels decreased from 2 mg/L in February 1989 to 0.9 mg/L in May.

The VT2-BNR process for operating two oxidation ditches in series with a side-stream anaerobic zone was put in service in June 1989 (Phase V). The process was developed after analyzing the limitations of the oxidation ditch system. It uses a side stream anaerobic zone which receives between 40 and 60 percent of the raw influent flow, and a mixed liquor stream flow between 30 and 45 percent of the raw influent flow (Figure 1). Using this arrangement, the plant was able to maintain monthly average ammonia-N levels of 1 mg/L and total nitrogen levels between 2 and 5 mg/L over a one year period of operation. The effluent total phosphorus averaged less than 0.6 mg/L during summer and 0.7 mg/L during winter. The plant performance for phosphorus continues to improve. Effluent average for May 1990 was 0.4 mg/L and should improve to 0.3 mg/L after the final upgrade is completed. The plant does not filter its secondary effluent. Filtration can decrease the effluent total phosphorus to 80 percent of observed levels.

The research attempted to understand some of the principles behind biological nutrient removal in oxidation ditches to develop process design

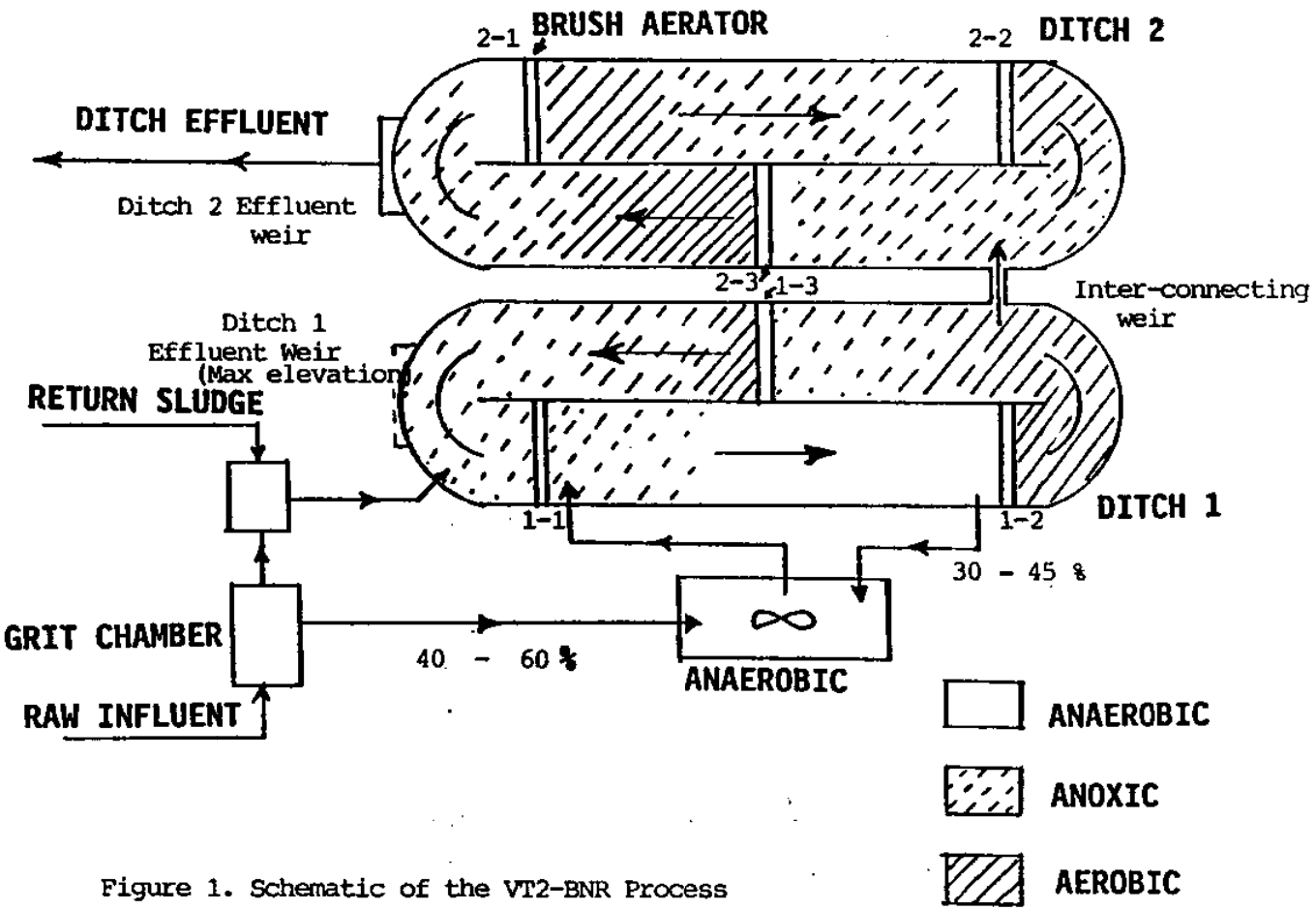


Figure 1. Schematic of the VT2-BNR Process

and operation criteria. The results from Bowie and other oxidation ditches have shown that they can be designed to achieve the performance levels similar to those attained in Phase V using one ditch with sufficient length (or liquid circulation time) and volume, and a side stream anaerobic zone receiving less than 50 percent of the influent flow.

Annual savings were calculated from the annual costs for chemical phosphorus removal, the cost to supplement the influent alkalinity with caustic soda, and the savings generated in aeration energy and sludge production as a result of excellent denitrification. For treating 2.2 MGD (8,300 m^3/day) of flow, $67,500 of savings are generated because of biological phosphorus removal alone. These are distributed between costs for chemicals for phosphorus removal (Ferrous Sulfate, $30,000) and supplemental alkalinity added as caustic soda ($37,500). Additional reduction in supplemental alkalinity requirements and aeration is generated by excellent denitrification. The effluent nitrate-N averaged less than 2 mg/L. This is approximately 4 mg/L less than the concentrations observed after modifications to activated sludge tanks using high rate processes. The alkalinity recovered during the denitrification of an additional 4 mg/L of nitrate-N reduced the cost for supplemental alkalinity by an additional $5,000. Net saving in energy for aeration, recycles and solids handling is $7,500 each year. By utilizing a BNR process with excellent nitrogen and phosphorus removal, the Bowie WWTP is saving $80,000 annually, when compared to a facility which may achieve the same degree of nitrification and phosphorus removal through chemical precipitation and is discharging 6 mg/L of nitrate-N in the effluent. Other than the cost for chemicals for phosphorus removal ($30,000), individual items are not reflected in a comparison of operating budgets for 1987 in 1990 because the plant discharged in excess of 10 mg/L of ammonia-N on an annual average prior to the modification. Nitrification increases aeration energy costs and alkalinity requirements substantially. This is offset partially by denitrification and biological P removal.

The laboratory staffing will have to be increased by 10 hours each week to conduct certain tests which are essential for satisfying the NPDES permit requirements for nitrogen removal. Operator training and certification

levels will have to be improved. These will result in expenditures of $23,000 each year. The return on this investment will be from an increase in the process efficiency.

The total volume available in the oxidation ditches is adequate for achieving biological nutrient removal to satisfy permits of 6 mg/L for Total Nitrogen and 1 mg/L for Total Phosphorus at the design flow of 3 MGD. The nominal HRT will be 24 hours at design flow. However, the aeration, clarifier solids handling and solids dewatering capacities will have to be increased to trea average flows above 2.5 MGD.

NUTRIENT REMOVAL PROCESSES IN OXIDATION DITCHES

Nitrogen Removal Processes

Oxidation ditches have been used for BOD removal since the early 1970s. Because they are constructed with nominal hydraulic retention times in excess of 10 hours in most cases, they can maintain MCRTs which are adequate for nitrification in warm weather without exceeding the solids handling capacities of the clarifiers. When an adequate amount of air is supplied, the nitrifiers will oxidize the ammonia to nitrates.

Towards the end of the 1970s, attempts were underway to improve nitrogen removal in oxidation ditches through biological nitrification and denitrification. The key to improving nitrogen removal was proper control of dissolved oxygen levels in different sections of the oxidation ditch while maintaining adequate mass of bacteria under aerobic and anoxic conditions. An anoxic condition is defined as an environment where the bacteria have nitrate or nitrite but have no dissolved oxygen available as electron acceptors for respiration. Several processes were developed for nitrogen removal, some of which are patented in the United States (Table 1). The oxidation ditches were aerated by a variety of equipment which included brush aerators, diffusers and submerged propeller devices and carousel aerators.

Enhanced Biological Phosphorus Removal Processes

Attempts to improve nitrification and denitrification in oxidation ditches led to some observations of increased biological phosphorus removal. However, in a majority of the cases, the capacity for biological phosphorus removal was not comparable to those observed in modified activated sludge systems. Examples of modified oxidation ditch systems where P removal was

Table 1. Oxidation Ditch Processes

Type	Manufacturer	Removal	Modifications
Carousels	Envirotech	BOD, Nitr	Denitrification
	Dwars	BOD, Nitr	
	Heederick and Verhay, Ltd	BOD, Nitr	
	Activox	BOD, Nitr	
Jet aeration channel	Pentech Houdaille	BOD, Nitr	
	Fluidyne Corp		
Orbal Processes	Envirex	BOD, N remv	Bio-P removal
Draft Tube Aerators (Barrier Oxid Ditches)	Inova Tech	BOD, N remv	
Rotating Diffuser Arm CLR	Schreiber	BOD, N remv	Bio-P removal
BioDenipho (anaerobic cell + SBR oxid. ditches)	Kruger	BOD, N, P removal	
Carousel + Anearobic + reair tanks	Eimco - modified Bardenpho	BOD, Nitr	N removal and bio-P removal with additional tanks
CLRs with single loops		BOD, Nitr	N removal Bio-P removal

CLR = continuous loop reactors
SBR = sequencing batch reactor
Nitr = Nitrification
N Removal = Biological Nitrification and Denitrification using influent or endogenous carbon sources

enhanced through accumulation of phosphorus in volutin granules (luxury uptake) include the concentric three ring ORBAL^{TM} system developed by ENVIREX and the BIODENIPHO system developed by Kruger Corp. The BIODENIPHO system uses an anaerobic cell in series with two or four oxidation ditches to enhance biological phosphorus removal. Multiple ditches are used to allow each ditch to operate as batch tanks which can be switched between aerated and non-aerated modes after certain intervals of time.

Biochemical Mechanisms

Biochemical mechanisms include reactions for BOD storage and removal under anaerobic, anoxic and aerobic conditions and reactions for nitrification. Several references are available in the literature on the mechanisms for

1) BOD storage under anaerobic conditions using the energy released during the breakdown of volutin (poly-P) granules;

2) Stabilization of BOD under anoxic conditions utilizing nitrates and nitrites as electron acceptors;

3) Stabilization of BOD under aerobic conditions with phosphorus uptake and formation of poly-phosphate granules, thereby increasing the phosphorus content in the biomass and the amount of phosphorus removed from the system in the biological sludge; and

4) Nitrification by autotrophic bacteria (Nitrosomonas and Nitrobacter) which respectively carry out a two step conversion of ammonia to nitrites and then to nitrates. The reactions occur under aerobic conditions and with the consumption of alkalinity as a carbon source.

A biochemical model for enhanced biological phosphorus uptake is discussed in Wentzel et.al. (1986). Mandt and Bell (1982) have discussed the stoichiometry of biological nitrogen removal.

PLANT AND PROCESS DESCRIPTION

The Bowie WWTP is an oxidation ditch system which receives 2.2 MGD of flow. The design capacity of the plant is 3 MGD. However, the WWTP receives raw influent whose BOD_5 concentrations are 50 percent in excess of those received by several other wastewater treatment facilities in the vicinity. The design capacity of the plant, expressed in terms of influent loadings, has not been included in the records. The capacity of the aerators, clarifiers and the solids dewatering units are not adequate if average flows exceed 2.5 MGD with BOD levels of 175 mg/L and COD levels of 450 mg/L.

Plant Description and VT2-BNR Process Flow Scheme

The plant receives wastewater flows from two separate sewer lines at its raw influent junction box. One line is pumped through force mains. The second line flows under gravity. The raw influent flow mixes with the recycle from the WWTP's solids handling processes at the junction box. The raw influent flow is not metered (Figure 2).

The "combined influent" (raw influent mixed with the plant recycle) is discharged into a flow equalization tank. In the original design, the flow equalization tank was operated with two separate tanks containing 315,000 gallons each. One of these tanks was modified to create an anaerobic cell. Each flow equalization tank has a slow speed mixer to keep the solids in suspension. Diffusers are installed in each tank to introduce air to increase the turbulence and avoid septic conditions.

Three variable speed pumps, each capable of pumping 3.3 MGD, pump the wastewater from the equalization tank(s) to the bar screens and the grit chamber. The pump speeds are set to pump between 2.5 and 3.0 MGD. Under

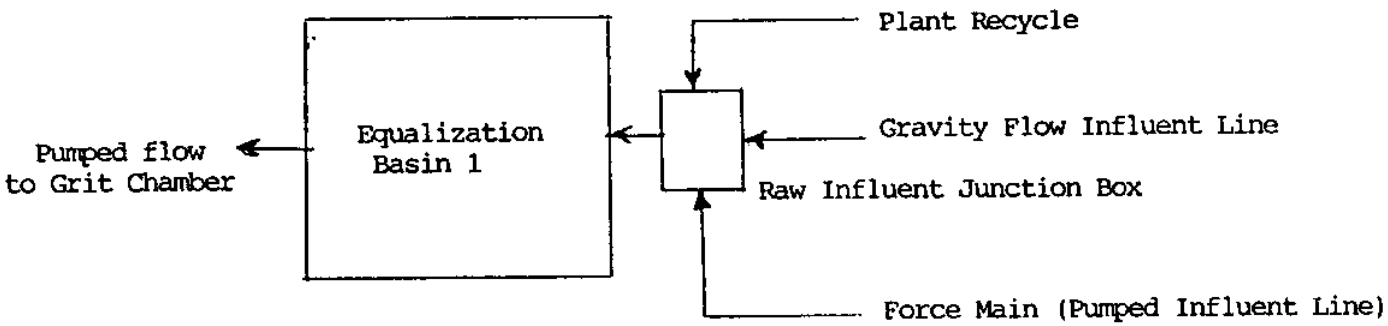


Figure 2. Schematic of Raw Influent Junction Box

normal conditions, one pump can handle all the flow received at the facility. The pumps are controlled by a pressure transducer which shuts them down when the liquid level in the equalization tank drops below a minimum set point.

A mechanical bar screen with 5/8 inch openings removes some of the fibrous material and plastics in the influent. The flow travels through a grit chamber where the organic material is kept in suspension with air bubbles introduced by diffusers. Only one half of the grit chamber was used after the modification to the VT2-BNR system. The air supplied by the diffusers was curtailed as much as possible.

The flow equalization tank adjacent to oxidation ditch 1 was modified to create an anaerobic zone. A gravity flow line was installed to carry up to 2000 GPM of flow from the grit chamber to the anaerobic cell. A second gravity flow line with a valve and metering device was installed to carry mixed liquor from oxidation ditch 1 to the anaerobic cell. The volume of liquid which can be stored in the anaerobic cell is 300,000 gallons, approximately.

Two submersible pumps were installed in the anaerobic cell to pump mixed liquor to oxidation ditch 1. Each pump is capable of handling 1100 GPM. The pumps will be upgraded to handle 1500 GPM, each, as part of the Phase VI modifications.

To operate the facility for biological phosphorus removal using the BNR process arrangement, it was determined that only 40 percent of the influent flow had to be sent to the anaerobic tank. The rest of the influent entered oxidation ditch 1 after it had mixed with the return sludge from the clarifiers at the flow splitter box (Figure 1).

Oxidation Ditch

Each oxidation ditch at the Bowie WWTP has a flow channel which is 660 feet long and 34 feet wide, and can hold 1.65 million gallons of liquid. Three brush aerators are available in each oxidation ditch and are spaced 220

feet apart. At an average velocity of 1 fps inside the oxidation ditch, the liquid circulation time is 11 minutes.

The flow splitter box can distribute the flow between the two ditches. In the VT2-BNR process arrangement, the two ditches were connected in series by cutting a 10 foot wide notch in the wall between them (Figure 1). The return sludge and influent enter oxidation ditch 1 in which partial BOD and nitrogen removal are achieved. The flow travels to ditch 2 over the inter-connecting weir. The effluent weir in ditch 1 (Figure 1) is raised to its maximum elevation to minimize the overflow from ditch 1 to the clarifiers, which would short circuit ditch 2 in the process.

To control aeration, the liquid level in the oxidation ditches can be adjusted by altering the level of the effluent weirs. This changes the depth of submergence of the brush aerators. Each brush aerator can also be operated on a timer. However, the RPM of each shaft and the number of brushes on the shafts are fixed.

The liquid level in ditch 2 is adjusted with its effluent weir to control the amount of air supplied in ditch 2. Indirectly, this controls the level of liquid in ditch 1 because the inter-connecting weir which connects the two ditches is fully submerged (that is, below the level of liquid in both ditches).

Clarifier and Solids Handling

The two clarifiers are rim-feed as shown in Figure 3. The effluent from ditch 2 is distributed equally between the two clarifiers. Each clarifier has a diameter of 65 feet, side water depth of 11 feet and a volume of 273,000 gallons. For an influent flow of 2.0 MGD, the nominal detention time in the clarifiers is 6.5 hours.

Analysis of loadings proved that the capacity of the secondary clarifiers is not adequate for the flows and solids loadings handled. During routine operation, the combined influent is pumped at a rate of 2.8 to 3.1 MGD from the flow equalization tank to the oxidation ditch. The plant operates at

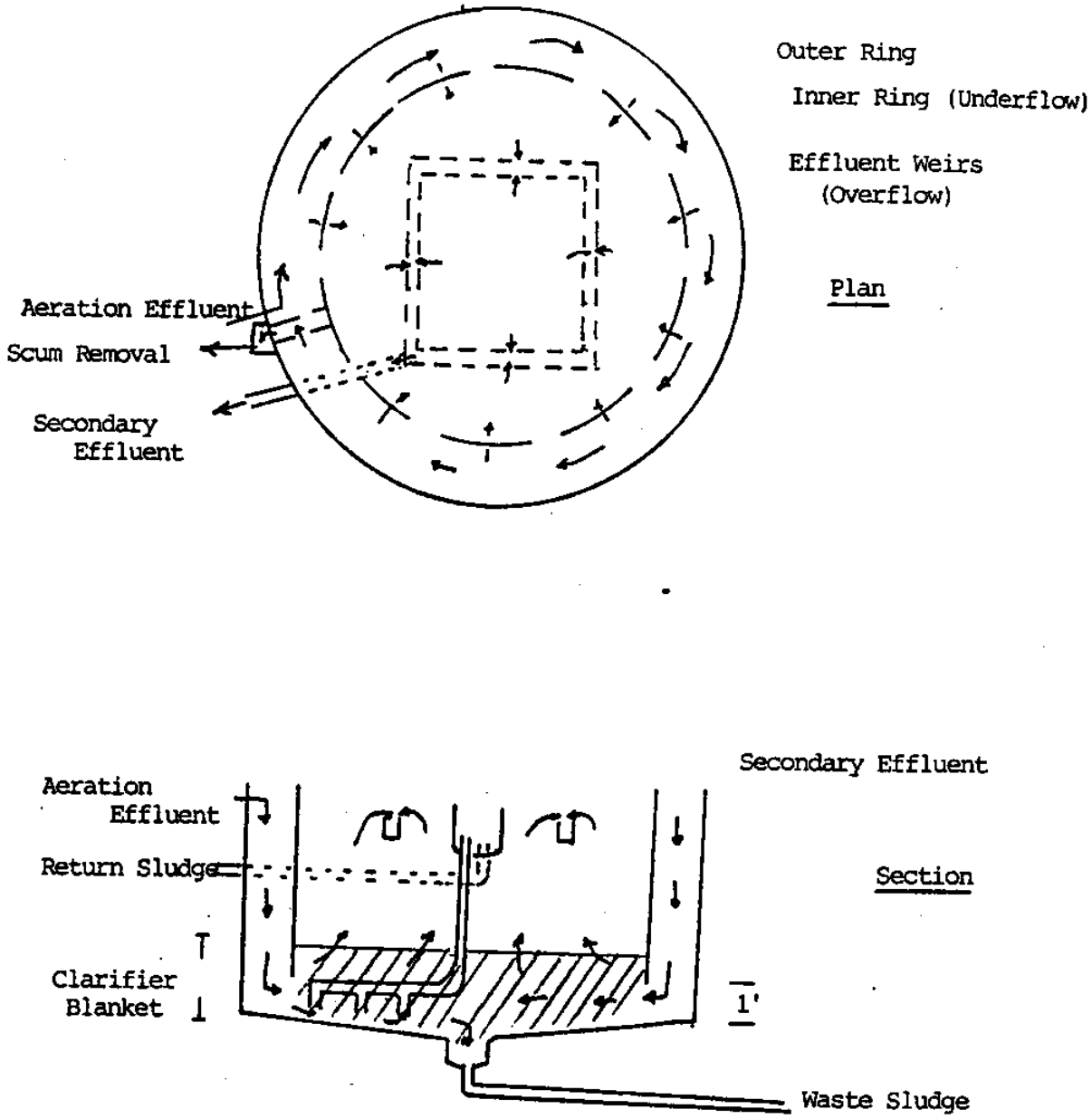


Figure 3. Schematic of Secondary Clarifier

MLSS concentrations between 2500 and 4000 mg/L. The surface overflow and solids loading rates are very high at the flows and MLSS concentrations at which the plant is operated (Table 2). It is necessary to maintain a small dose of polymer to enhance sludge settling rates and increase the capacity of the clarifiers.

The return sludge is siphoned through a travelling rake arm off the bottom of the clarifier. The waste sludge is drawn off a hopper at the bottom of the clarifier (Figure 3). The return sludge is pumped back to the flow splitter box. A temporary return sludge chlorination line was installed to inject a maintenance dose of chlorine (less than 0.5 pounds per thousand pounds of volatile solids) to control the population of filamentous bacteria.

The waste sludge is pumped from the clarifiers to gravity thickener # 2. Until August, 1989, the thickened sludge from the bottom of this thickener was purifaxed and stored in gravity thickener # 1 before dewatering on the belt filter presses. The purifax system was taken out of service in September 1989 and a lime stabilization unit was installed.

The plant has two belt filter presses. Their dewatering capacity controls the minimum operating MCRT. When the belt presses are operated for 8 hours each day and for 5 days each week, it is difficult to maintain MCRTs less than 20 days. Up until the end of 1988, the second oxidation ditch was used as an aerobic digester to reduce the solids loading on the clarifiers and the filter press and maintain MCRTs of 10 to 15 days. The presses were operated over two 8 hour shifts each day when sludge was not wasted to the aerobic digester. However, with two ditches in service since February 1989, the WWTP has operated at MCRTs between 20 and 35 days and has eliminated the second shift on the filter press.

The washings from the filter press and the overflow from the thickeners and scum handling tanks are returned to a plant drain tank. These flows are pumped to the raw influent junction box.

Table 2. Actual and Recommended Secondary Clarifier Loadings

	Actual	Recommended Maximum[1]
No of secondary clarifiers	2	
Outer Ring Diameter	65 feet	
Side Water Depth	10 feet	
Surface Area per clarifier	3318 ft^2/ clarifier	
Flow to clarifier	3.0 MGD Influent 2.0 MGD Return	
Surface Overflow rate at 3.0 MGD	452 gpd/ft^2	400 gpd/ft^2
Length of Weir per clarifier	367 feet	
Weir overflow rate	4079 gal/ft	
Solids Loading Rate at 3500 mg/L	22 lb/ft^2	20 lb/ft^2

[1] Manual of Practice, 9, WPCF.

It is recommended that clarifiers be designed for solids loading rates of 15 lb/ft^2. The plant has to operate at MLSS levels below 2400 mg/L to satisfy this criterion.

Secondary Effluent Treatment

The secondary effluent is chlorinated with chlorine gas for disinfection. Sodium metabisulfite is added for dechlorination. The effluent is reaerated before discharge. A 24 hour composite sampler is used to sample the effluent at 15 minute intervals of time.

Chemical Addition

Caustic Soda

Caustic soda was added to supplement the influent alkalinity between September 1988 and May 1989. For some period of time, the caustic soda was added directly to the oxidation ditches. At a later date, the caustic soda was added to the plant drain tank, mixed with the recycle from the solids handling units, and pumped back to the raw influent junction box. The caustic soda solution was reduced from a 50 percent solution to 30 percent in winter.

Discontinuation of chemical P removal, increased reliability and efficiency of denitrification, and the reduction in the chlorine dose necessary for filamentous bacteria control have helped reduce the alkalinity consumption and increase the pH in the biological system. Since June 1989, the WWTP has been able to eliminate supplemental alkalinity requirements.

Ferrous Sulfate

Ferrous sulfate was added to remove phosphorus chemically (and supplement excess biological uptake of phosphorus - as may have been the case at the Bowie WWTP during Phases I to III) and satisfy a monthly permit of 1 mg/L for total phosphorus. It was discontinued after August 17, 1988, to initiate experiments which would examine the extent of biological phosphorus removal possible in single loop oxidation ditches.

Polymer

An anionic polymer was added to the oxidation ditch effluent to enhance sludge settling rates in the clarifiers when ferrous sulfate was added to remove phosphorus. A cationic polymer was used after ferrous sulfate addition was discontinued.

A cationic polymer is used for solids dewatering.

Sampling and Analysis

Sampling

Raw Influent

The raw influent was sampled at the junction box (Figure 2). Several attempts to determine the true concentration of the raw influent by separately measuring the strength of the "combined influent" (raw influent mixed with the plant recycle from solids handling units) and the plant recycle, and developing a mass balance using the flows and concentrations failed to yield satisfactory results. In January 1989, a procedure was developed to collect the raw influent from the junction box, ten minutes after switching the plant recycle pumps off. During the phases when the WWTP was manned through two shifts each day, a composite sample was collected by sampling the raw influent three times over a sixteen hour period. After the plant switched to operation with one shift only, the raw influent was composited by collecting three samples within an eight hour period.

Combined Influent (Raw Influent mixed with Plant Recycle from Solids Handling)

A 24 hour composite sampler was used to collect a sample of the combined influent after it had passed through the bar screen and before it entered the grit chamber.

Field measurements for dissolved oxygen, pH, alkalinity and temperature were made at the grit chamber.

Anaerobic Cell Sample

A sample of the mixed liquor from the anaerobic cell was collected several feet away from the points where flows entered the anaerobic cell to ensure a representative sample of the mixed tank. Field measurements for D.O., pH and oxidation reduction potentials were conducted on samples collected near the center of the basin.

Oxidation Ditch 1

Samples were collected in ditch 1 at the point marked Y in Figure 4 for analysis at the plant and the Occoquan Watershed Lab. This location was selected because

i) it was readily accessible, and
ii) concentrations of most parameters of interest were fairly uniform along the length of the ditch (see below).

Ammonia, nitrate, ortho-phosphorus and soluble COD levels were relatively constant across the length of the ditch because the internal recycle rate was about 90 times the influent flow (2.2 MGD). The ratio of the flow rate across a cross-section of an oxidation ditch to the influent flow is a measure of the internal recycle rate. At Bowie, the flow path in the oxidation ditch is 30 feet wide and 10 feet deep. When the velocity of liquid is 1 fps, the flow rate across a section of the ditch is 194 MGD.

During phases II and III, only one oxidation ditch was used for secondary treatment. Samples were also collected prior to brush aerator 1-2 (Figure 4). On several occasions, dissolved oxygen levels were measured at several points to obtain a profile along the length of each ditch. The measurements were made between one and five feet below the water surface.

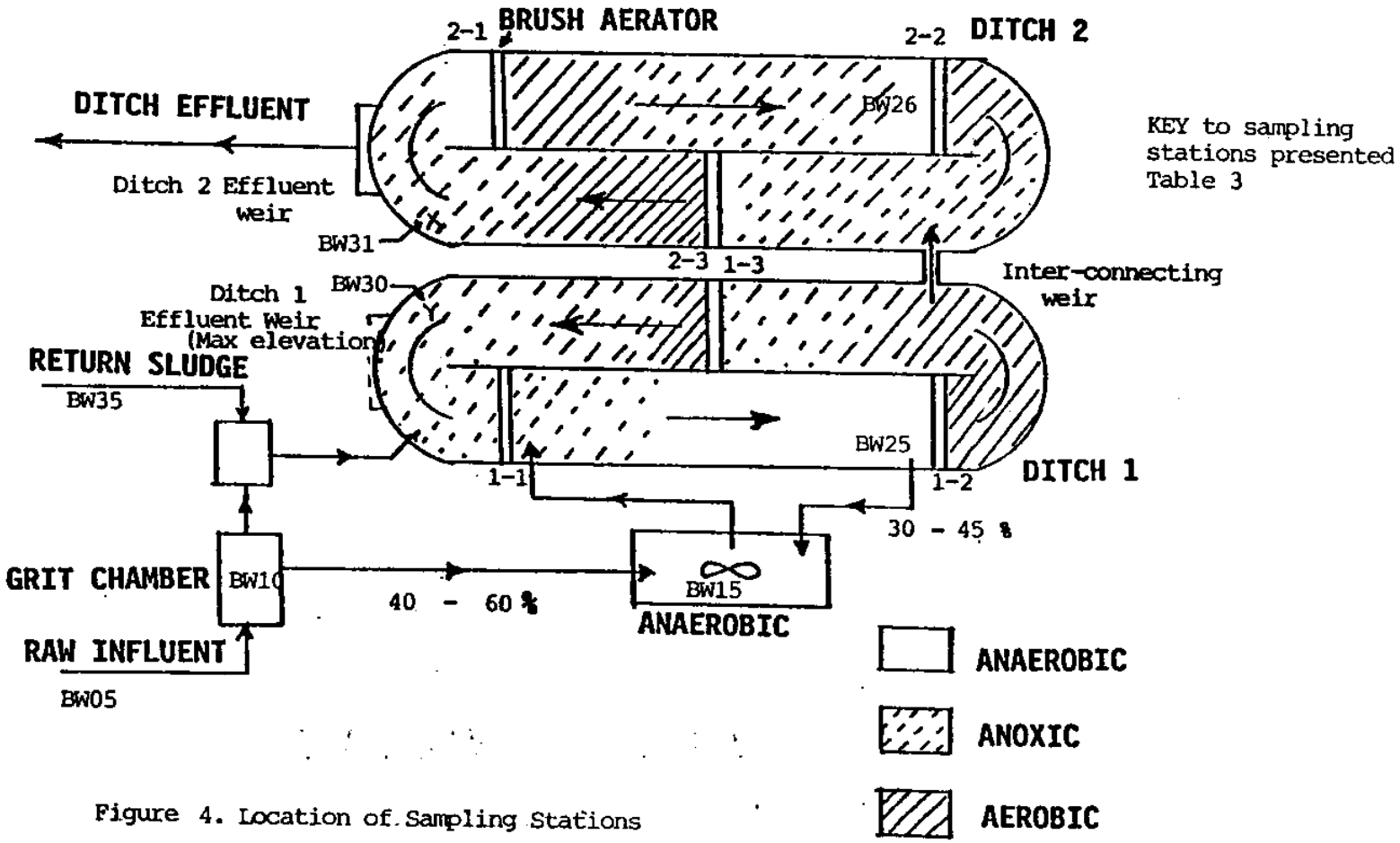


Figure 4. Location of Sampling Stations

Table 3. Sampling Stations

Station Code	Description
BW05	Raw Influent
BW10	Combined Influent (Raw Influent mixed with Plant Recycle)
BW15	Anaerobic Cell
BW25	Oxidation Ditch 1, prior to brush aerator 1-2
BW30	Oxidation Ditch 1, 100 feet after brush aerator 1-3
BW26	Oxidation Ditch 2, prior to brush aerator 2-2
BW31	Oxidation Ditch 2, prior to effluent weir
BW35	Return Sludge
BW40	Waste Sludge
BW50	Secondary Clarifier Effluent
BW55	Gravity Thickener Effluent
BW60	Filter Press Influent
BW62	Belt Filtrate
BW65	Plant Recycle (effluent from all solids handling units)
BW70	Final Effluent Composite after chlor. and dechlorination
BW71	Final Effluent Grab after chlorination and dechlorination

Oxidation Ditch 2

Samples were collected in ditch 2 at the point marked as X in Figure 4. This location was within a few feet of the effluent weir. On several occasions, dissolved oxygen profiles were measured along the length of this ditch.

Secondary Effluent

Secondary effluent samples were collected at the plant to conduct process control tests for alkalinity, turbidity and ortho-phosphorus.

Final Effluent

A 24 hour composite sample of the final effluent was collected after chlorination and dechlorination. The composite sample was not flow weighted.

Waste Sludge

The waste activated sludge was sampled on several occasions. However, because of the location of the sampling port on the wall of a straight length of pipe instead of a bend, wall effects caused an error in the MLSS of the sample collected. The wall effect results in higher solids concentrations near the center of the pipe cross-section and lower solids concentrations near the inner surface.

Gravity Thickener Supernatant

The gravity thickener supernatant was sampled to determine the extent of phosphorus released from the biomass. Nitrogen and COD concentrations were also measured.

Belt Filtrate

The washings from the filter press were sampled on a few occasions. The concentration of COD and nutrients in the sample were a function of the efficiency of the filter press at the loading experienced at the time of collection.

Plant Recycle

Several composite samples of the recycle from the solids handling units to the headworks were collected and analyzed. The composite samples of the plant effluent had a wide range of values, possibly because of the nature of the processes involved.

The sampling stations and the corresponding codes are presented in Table 3.

Analysis

Dissolved Oxygen

Dissolved Oxygen measurements were conducted with an YSI D.O. probe and meter. The measurements were made in situ to eliminate errors which may be observed because of reaeration of, or oxygen uptake from, the samples after they had been collected.

Oxidation Reduction Potential

Samples were collected in small bottles after taking precautions to avoid reaeration. A Platinum/Ag/AgCl Composite Electrode (Fisher 13-620-82) ORP probe was used with a SA 250 pH/mV/C meter to measure oxidation reduction potential. The probe was soaked in warm water for several minutes before its use to remove any coating of foreign material from the surface of the electrode. This allowed the reading to stabilize within a few minutes after it was introduced into the liquid.

pH

All pH readings were taken on site within three minutes after the collection of the sample.

Alkalinity

A Hach digital titrator was used for field measurements. Laboratory measurements were carried out with 0.02 N sulfuric acid and bromocreosol green indicator.

PROCESS OPERATION AND RESULTS

Phase I (1985 to August 1987)

During Phase I, the plant was operated with one oxidation ditch to treat flows of 2.2 MGD. Two brush aerators were operated over a 24 hour period and the third aerator was operated for 12 hours each day. To adjust for the increase in oxygen transfer into the liquid phase at lower temperatures, the level of the liquid in the oxidation ditch was lowered each winter to reduce the depth of submergence of the brush aerators. The operators had also observed that the pH levels in the effluent dropped close to the permitted minimum of 6.5 when three brush aerators were operated continuously. By limiting the air transferred by the third brush aerator -operating it over a fraction of the day with a timer - they were able to curtail nitrification and increase denitrification. This reduced the alkalinity consumed during nitrification and increased the alkalinity recovered during denitrification. Though a caustic soda tank was installed in 1985, caustic soda was not used to supplement the influent alkalinity.

Since July 1986, phosphorus was removed chemically by adding ferrous sulfate to satisfy a permit of 1 mg/L for total phosphorus. The plant was able to operate with about 50 percent of the recommended dose for ferrous sulfate because anaerobic conditions near the bottom of the oxidation ditch created conditions which enhanced biological phosphorus removal. Iron was added at a rate of 10.9 mg/L (as Fe^{2+}) to the combined influent (raw influent + plant recycle).

It had been observed that it was difficult to control the depth of the sludge blanket in the clarifiers at MLSS levels above 3000 mg/L. Because of the arrangement of the rim feed clarifiers, where the flow is introduced along the rim and one foot off the bottom, increasing the return sludge

flow rate did not enhance the sludge settling rates. The hydraulics are discussed the following chapter. To maintain MLSS levels at 3000 mg/L and operate within the capacity of the gravity thickeners and the belt filter presses, a fraction of the biomass had to be discharged periodically to the second oxidation ditch. The biomass was digested aerobically in the second oxidation ditch. This reduced sludge handling and disposal expenses because it reduced the sludge production. However, it increased aeration energy costs because one or more brush aerators had to be operated in the second oxidation ditch to digest the sludge.

In the first few years of operation, the plant had faced a number of problems with the brush aerators. However, the initial problems were gradually eliminated after new shafts were fabricated and a maintenance program was implemented. The reliability of the brush aerators was a key factor in assessing the options available for operating the plant with one oxidation ditch only.

The plant did not have a return sludge chlorination line to control filamentous bacteria. The Sludge Volume Index (SVI) averaged between 100 and 230 mL/g (monthly average). The SVIs were the lowest in winter and the highest between May and October (Figure 5). In warm weather, the extent of nitrification was limited by the oxygen supply. Efficient scum removal along the rim of the clarifiers helped keep the problems with foam and Nocardia, a type of filamentous mirco-organism, in check (Figure 3).

Phase II (September 1987 to July 1988)

During this phase, the plant was operated with one oxidation ditch only. The results are presented in Tables 4 to 7.

Between October 1987 and May 15, 1988, the method of aeration was modified to inject most of the air in one section of the ditch. Analysis completed towards the end of Phase II showed that the plant did not have sufficient aeration capacity to nitrify with two brush aerators only.

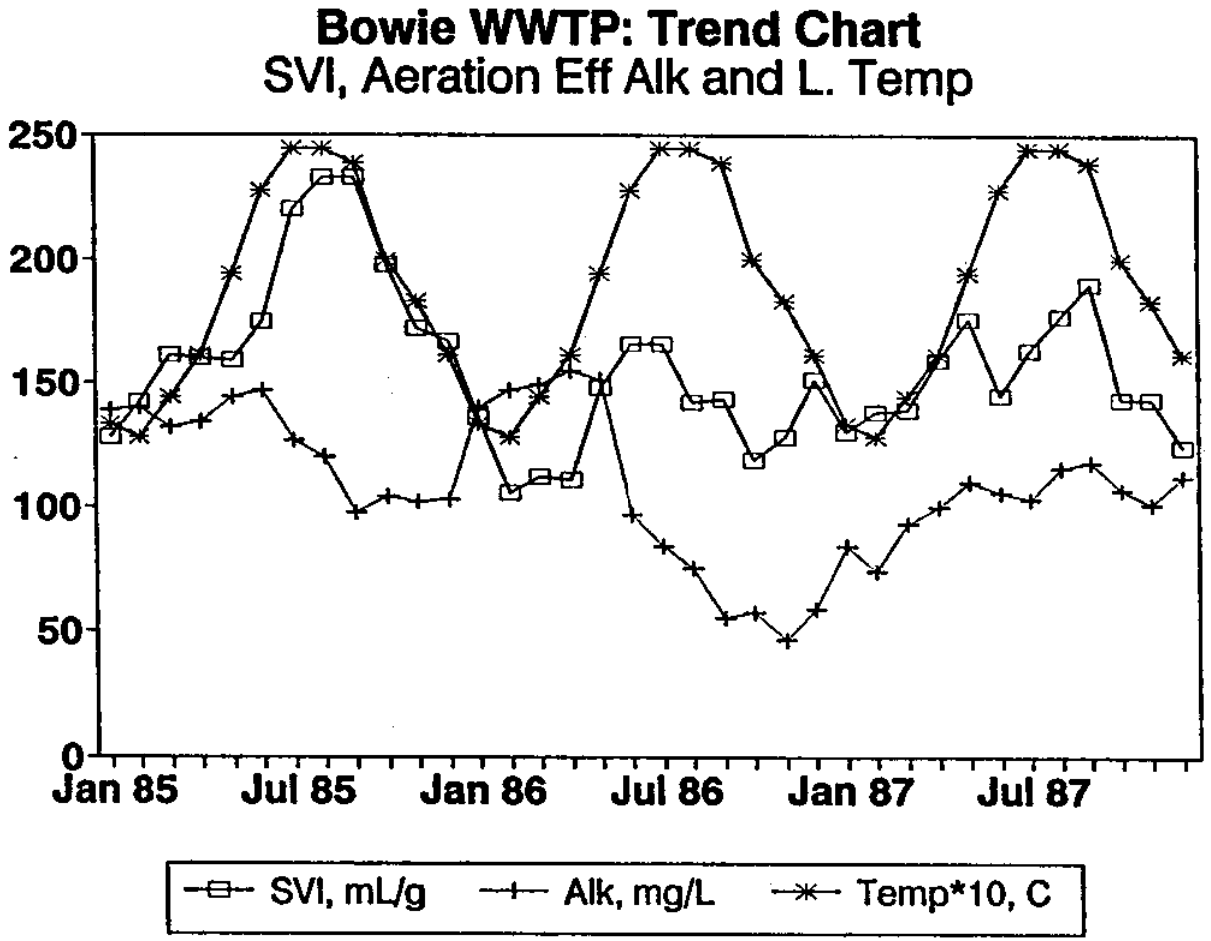


Figure 5. Trend Chart for SVI, Alkalinity and Liquid Temperature in Phase I

Table 4. Operating Data for Oxidation Ditch System, Bowie WWTP

	CFlow MGD	FFlow MGD	WASF 1000GP	AnaEn KWH	RotEn KWH	AtmT F	Tb NTU	NaOH GPD	SVI mL/g	Dry Solids lbs/day		AlkD1 mg/L	AlkD2 mg/L	SSD1 mg/L	SSD2 mg/L	AnaFl MGD	MLRFl MGD	WaterT F
Sep 87							3.6		190	3341		118		2867				
Oct 87							2.9		169	4349		107		3615				
Nov 87							3.5		143	3283		101		2676				
Dec 87							5.7		124	2259		112		2008				
Jan 88							6		121	3933		110		2733				
Feb 88	2.19						8		101	4515		113		2830				
Mar 88	2.21		37.5				8.6		103	4469		130		2918				
Apr 88	2.25		28				13.3		119	4765		123		2517				57
May 88	2.37		27.2				4.7		112	4069		114		3166				60
Jun 88	2.34		17.8				6.6		128	5021		71		3969				60
Jul 88	2.55		42.1		2672	93.2			174	4440		71		3726				66
Aug 88	2.56		27.8		2769	89.6			221	4983		71		3099				75
Sep 88	2.54				2481	77.9			164	3776		94		2574				78
Oct 88	2.15				2100	59	5.5		143	2409		106		2923				74
Nov 88					2210	57.2	5.6		138	2897		119		3664				69
Dec 88					1800	45	4.3		154	3442		96		3539				66
Jan 89	2.66		25.7		1610	44			143	3590		95						61
Feb 89	2.8				1843	42		28.6	165	4422		90	84	3173	3072			58
Mar 89		2.125	44.3		1958	52	3.5	33.3	142	5337		81	83	2760	2741			57
Apr 89	2.62	2.104	24.6	150	2117	63.2	2.9	7.4	131	4379		79	80	2146	2441	0.4	1	56
May 89	2.65	2.419	23.1	186	2167	70	2	39.5	114	4161		76	77	2210	2576	0.6	1	62
Jun 89	2.73	2.513	33.25	290	1600	89.6	1.9	2.4	134	5147		84	87	2784	2996	1.8	1.5	64
Jul 89	2.68	2.15	25.4	345	1844	86.7	1.7	0	147	4743		88	72	2708	2666	1.8	1.44	73
Aug 89	2.47	1.96		337	1982		1.3	0	151	2799		95	73		2503	1.8	1.44	74
Sep 89	2.42	1.97	30.1	0	1994	78.4	1.5	0	177	3327		101	72		3032	0	0	76
Oct 89	2.08	1.95	27.8	323	2075	68.8	1.6	0	185	2440		97	72		2940	1.76	1.21	72
Nov 89	1.97	1.86	25.5	321	1965	50.2	2	0	155	3459		97	73		3328	1.73	1.11	69
Dec 89	2.21	1.96	28.1	317.1	1991	30.6	1.6	0	155	3137		99	75		3690	1.53	1.07	63
Jan 90	2.42	2.06	26	308	1983	47.9	1.5	0	148	3030		93	75		4042	1.34	1.07	54
Feb 90		1.99	25.5	302	1890	51	1.9	0	124	3973		94	75		3950	1.35	1.08	56
Mar 90	2.05	1.97	29.1	336	2076	54.2	2.1	0	128	3975		98	73		4113	1.38	1.07	57
Apr 90		2.07		346	1963		2.1	0	116	4364		101	77		3427	1.39	1.19	60
May 90		2.11		380	2200				119									
Jun 90				390	2236													
Ph II	2.318	ERR	30.52	ERR	2672	93.2	6.589	ERR	129.4	4110	ERR	105.2	ERR	3016	ERR	ERR	ERR	60.75
Ph III	2.45	ERR	25.7	ERR	1930	56.62	5.133	ERR	148.4	3223	ERR	102	ERR	3175	ERR	ERR	ERR	69.6
Ph IV	2.69	2.216	30.67	168	2021	56.8	2.8	27.2	138	4575	ERR	81.5	81	2572	2708	0.5	1	58.25
Ph V	2.337	2.047	27.86	300.4	1964	61.93	1.745	0.218	144.9	3672	ERR	95.18	74.91	2746	3335	1.444	1.107	65.27

Please Refer to Appendix A for Complete Key to Column Headings.

Monthly Average Values are shown to the left.

Phase Averages

Table 5. Operating Data for Anaerobic Cell and Ditch 1 (Before 1-2), Bowie WWTP

	BW15 Anaerobic Cell											BW25 Ditch 1 Before Aerator 1-2							
	D.O.	ORP	pH	OP	NH3	NO2	OxN	SCO	COD	TSS	VSS	DO	ORP	OP	NH3	OxN	SCO	TSS	VSS
	mg/L	mv		mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L	mv	mg/L	mg/L	mg/L	mg/L	mg/L	mg/L
Sep 8																			
Oct 8																			
Nov 8																			
Dec 8																			
Jan 8																			
Feb 8																			
Mar 8													112					2918	2200
Apr 8												0.42	122	0.27		0.02		2560	1936
May 8												1.42	168	0.6		1.05		3209	2265
Jun 8												0.94	191	0.41	0.27	2.83			
Jul 88												0.25		0.06	1.9	1.04		3726	2665
Aug 8												0.22		0.42	1.58	1.06		3099	2409
Sep 8												0.25		0.86	4.64	0.08		2828	2210
Oct 8												0.23		0.25	1.99	0.72		2637	2080
Nov 8																			
Dec 8																			
Jan 8												1.55		0.07	10.4	0.16		3800	3200
Feb 8																			
Mar 8																			
Apr 8	0.01		6.9	3.65	4.24	0.02	0.06	24.4	2732	1950	1507								
May 8	0.03		7	4.26	7.41	0.02	0.07	38.1	2027	1447	1132								
Jun 8	0.07		6.9	3.51	16.4	0.01	0.14	24	1824	1390	1095								
Jul 89	0		7.09	5.14	15.4	0.01	0.08	30.8		1340	1086								
Aug 8								58											
Sep 8												0	115	1.87	8.63	0.09		3302	2835
Oct 8	0.01	-72	7.2	7.75	17.5	0.01	0.03	55.9	1677	1434	1136								
Nov 8	0	-39	7.1	6.97	16.8	0.03	0.64	58.9	2030	1916	1523								
Dec 8	0.03	24	7.2	8.06	20.1	0.01	0.02	40.7	2411	2015	1615								
Jan 9	0.02	-27	7.2	8.93	17.6	0.01	0.04	48.6	2622	2217									
Feb 9	0	-75	7.3				0.03	45.7		2060	1680		-32	0.54	0.05	0.5		3910	
Mar 9	0.1	4	7.2																
Apr 9																			
May 9																			
Jun 9																			
Ph II	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	0.76	148	0.34	1.09	1.24	ERR	3103	2267
Ph III	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	0.68	ERR	0.39	5.68	0.32	ERR	3088	2497
Ph IV	0.02	ERR	6.95	3.96	5.83	0.02	0.07	31.3	2380	1699	1320	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR
Ph V	0.03	-31	7.15	6.73	17.3	0.01	0.14	45.3	2113	1767	1356		41.4	1.21	4.34	0.3	ERR	3606	2835

Table 6. Operating Data for Ditch 1 and Ditch 2, Bowie WWTP

	BW30 Ditch 1														BW31 Ditch 2												
	DO mg/L	ORP mv	pH	T C	OP mg/L	TP mg/L	NH3 mg/L	TKN mg/L	NO2 mg/L	OxN mg/L	SCO mg/L	COD mg/L	TSS mg/L	VSS mg/L	DO mg/L	ORP mv	pH	T C	OP mg/L	TP mg/L	NH3 mg/L	TKN mg/L	NO2 mg/L	OxN mg/L	SCO mg/L	COD mg/l	TSS mg/L	VSS mg/l
Sep 8																												
Oct 8																												
Nov 8													2676	1875														
Dec 8													2008	1547														
Jan 8																												
Feb 8																												
Mar 8	0.32	112	7.1	13.7	0.2	74.2		205	0.01	0.01			2918	2200														
Apr 8	0.55	122	6.9	15.6	0.13	65.6	0.2	176	0.01	0.04			2517	1869														
May 8	1.53			19																								
Jun 8	0.62			20.9	0.36					3.25																		
Jul 88	0.22		6.8	24.4	0.05	104	3.13	216	0.29	1.32	23.9	3866	3726	2665														
Aug 8	0.27		6.8	25.7	0.37	90.6	1.79	180		1.14	28.7	2988	3099	2408														
Sep 8	0.22		6.7	23.5	0.77	72	4.11	153	0.1	0.14	43.5	3173	2574															
Oct 8	0.25		6.8	21	0.22	93	2.05	202	0.46	0.72	39.4	4219	2923															
Nov 8																												
Dec 8																												
Jan 8	1.4			14.3	0.05	108	8.53	325	0.04	1.24	33.4	5639	4120															
Feb 8	0.3		6.9	13.8	0.72	119	1.13	211	0.19	0.87	19.1	4102	3648	2763	0.45		6.93	13.8	0.74		1.51			0.92	19.4		3170	2348
Mar 8	0.46		6.91	13.3	0.64	94	0.5	195	0.26	1.8	24.2	3665	2784	2120	0.46		6.9	13.3	0.79		1.19		0.23	1.22	29.8		2845	2172
Apr 8	0.26		6.85	15.9	0.71	73	0.24	152	0.61	1.72	25.9	2933	2215	1705	0.26		6.89	15.9	0.56	82.6	0.19	167		0.96	20.5		2510	1912
May 8	0.3		6.91	18.3	0.8	71	0.74	145	0.1	1.88	20.7	2863	2085	1587	0.3		6.93	18.3	0.67	82.6	0.65	163	0.04	1.22	20.4		2455	1880
Jun 8	0.03		6.93	22.5	0.18	83	5.35	173	0.05	0.72	16.5	3192	2590	1985	0.05		6.87	23.2	0.1	89.3	0.53	183	0.03	1.26	12		2740	2085
Jul 89	0		6.94	23.6	0.7	89	5.16	174	0.09	0.4	18		2505	1933	0.02		6.91	23.8	0.2	87.4	0.42	169	0.07	1.44	20.4		2546	2010
Aug 8											32														17			
Sep 8			7.1	23.1	1.64	127	8.5	214	0.01	0.08	21.6	3772	2920	2207	0.16	171	6.98	22.8	0.66	114	0.48	197	0.07	4.16	17.9		2960	2222
Oct 8	0.01		7.1	21.4	1.59	108	5.12	188		0.32	53.6	3457	2856	2178	0.8	40	7.1	21.3	0.58	121	0.08	211	0.03	3.49	30.2		2816	2136
Nov 8	0.01		7	18.5	2.1	132	6.24	228		0.37	34.9		3703	2860	0.83	82	6.95	18.1	0.5	130	0.23	216	0.07	3.86	18.4		3453	2620
Dec 8	0.05	54	7	13.8	1.68	137	5.7	254		0.68	22.6		3608	2825	0.84	116	7	13.4	0.45	147	1.54	265	0.04	3.61	19.7		3805	2960
Jan 9	0.12		7	14	0.82	152	5.04	294		1.05	20.8		4205		0.54	69	6.9	13.7	0.3	153	0.03	283	0.02	2.32	20.2		4173	3295
Feb 9															0.34	25	6.8			145	0.04	300			21.4			
Mar 9															0.6	35	7											
Apr 9																												
May 9																												
Jun 9																												
Ph II	0.65	117	6.93	18.7	0.19	81.3	1.67	199	0.1	1.16	23.9	3866	2769	2031	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR
Ph III	0.62	ERR	6.75	19.6	0.35	91	4.9	227	0.2	0.7	38.8	4344	3206	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR
Ph IV	0.33	ERR	6.89	15.3	0.72	89.3	0.65	176	0.29	1.57	22.5	3391	2683	2044	0.37	ERR	6.91	15.3	0.69	82.6	0.89	165	0.14	1.08	22.5	ERR	2745	2078
Ph V	0.04	54	7.01	19.6	1.24	118	5.87	218	0.05	0.52	27.5	3474	3198	2331	0.46	76.9	6.95	19.5	0.4	123	0.42	228	0.05	2.88	19.7	ERR	3213	2475

Table 7. Final Effluent Monthly Averages and other Operating Parameters, Bowie WWTP

	BW70	Final Effluent										Calculated Parameters									
	OP mg/L	TP mg/L	NH3 mg/L	TKN mg/L	NO2 mg/L	OxN mg/L	SCOD mg/L	COD mg/L	BOD mg/L	TSS mg/L	VSS mg/L	T C	Alk mg/L	TN mg/L	MCRT days	%P TSS basis	%P VSS	%N TSS basis	%N VSS	DN mg/L Inf flow basis	%Vol Solids
Sep 87	0.24	0.45	13.5	16.5		0		26	7					16.5	14.9						
Oct 87	0.22	0.47	9.83	12.8		0.2		29	26	8				13	10.8						
Nov 87	0.32	0.66	7.62	10.4		0.4		35.3	9.5	8				10.8	13.9						
Dec 87	0.5	0.8	17	20.8		0		65	20.3	22				20.8	14.9						
Jan 88		0.9	18	21		0.01		66	18	31				21.01	9.6						
Feb 88	0.34	1	19.5	21.6		0.01		69.3	16.8	28		12.8	141	21.61	10.2						
Mar 88	0.4	0.84	20.1	24.5	0.01	0.01		84.5	25	16		13.5	152	24.51	10.2	2.52			8.24		75.4
Apr 88	0.48	0.77	18.8	23.1	0.01	0.02		94.5	19.2	28		15.8	147	23.12	7.73	2.49			7.98		75.6
May 88	0.52	0.87	17.2	20.1	0.32	0.44		47.5	9.2	13		19.2	131	20.54	10.6	2.98			9.05		70.5
Jun 88	0.53	0.79	0.95	2.56	0.52	0.68		34	6	18		21.4	79	3.24	10.9	3.17			8.09		
Jul 88	0.34	0.67	0.51	1.83		1.72	33	42.3	14.6	10		24.9	69.8	3.55	11.5	2.79		6.07			71.5
Aug 88	1.31	1.83	0.56	2.17		1.31	28.8	38	13.8	8		25.7	72	3.48	8.6	2.92		5.79			77.7
Sep 88	1.91	2.24	2.69	4		3.01	36.9	48.7	22	5		23.4	101	7.01	10.3	2.81	3.57	5.96	7.56		78.8
Oct 88	1.53	1.79	1.21	2.71		1.82	30.1	35.1	16	4		20.7	99	4.53	16.6	3.17	4.02	6.92	8.77		78.9
Nov 88		1.51	0.18	1.6		1.77			27	22				3.37	12						
Dec 88		1.99	5.84	6.92		3.11			20	20				10.03	13						
Jan 89	1.61	2.25	4.99	6.52	0.17	3.77	36.4	53	9	12		14	125	10.29	13						
Feb 89	1.66	1.95	0.44	1.6	0.07	1.68	22.23	25.2	12	11		13	81	3.28	15		4.41				74
Mar 89	1.55	1.69	0.52	1.22	0.07	1.81	27.9	30.7	7	9.8		13	80	3.03	14.2		4.42				76
Apr 89	0.89	1.2	0.17	1.51		0.79	19.8	27.5	4	7.1			83	2.3	16.5		4.3				77
May 89	0.81	0.88	0.28	0.94		1.89	20.5	25	5	4.9			82	2.83	18		4.48				76
Jun 89	0.32	0.59	0.13	0.99		1.18	13.8	22	6	2.9		23.3	73	2.17	18	3.92	4.28				76
Jul 89	0.35	0.43	0.1	0.8		1.27	9.95	30.2	4	0.6		23.9	73	2.07	17	4.15	4.35				79
Aug 89		0.5	0.1	0.8		1.5			5	2				2.3	24.6						
Sep 89	0.79	0.97	0.1	0.68		2.98	12.4	25.6	3	3		23	80	3.66	27.6	3.85	5.14	6.66	8.87	18.2	75
Oct 89	0.71	0.9	0.06	0.82		3.55	19.85	29.7	3	6		20.8	75	4.37	36.5	4.33	5.7	7.51	9.91	27.6	75.6
Nov 89	0.54	0.63	0.18	0.93		3.9	14.6	18.4	3	7		17.7	72	4.83	29.1	3.76	4.96	6.27	8.26	19.7	75.9
Dec 89	0.47	0.57	1.1	1.2		2.4	17.44	19	3	6		13	80	3.6	35.6	3.86	4.97	6.97	8.97	21.2	77.8
Jan 90	0.61	0.64	0.1	0.92		2.77	17.5	18.4	3	3		13.5		3.69	40.3	3.73	4.66	7.01	8.68	19.6	79
Feb 90		0.55		1.1					3	3					30.1	3.99	4.98	8.26	10.3	5.1	80
Mar 90		0.78		0.7		3.5			3	3				4.2	31.2						
Apr 90		0.48		1.1		1.7			3	2				2.8	25.2						
May 90		0.35		1		1.5			3	2				2.5	21.1						
Jun 90																					
Ph II	0.406	0.777	12.95	15.87	0.215	0.349	33	56.74	16.46	18.2	ERR	17.93	120	16.22	11.03	2.79	ERR	6.07	8.34	ERR	73.25
Ph III	1.683	1.956	2.982	4.35	0.17	2.696	34.47	45.6	18.8	12.6	ERR	19.37	108.3	7.046	12.98	2.99	3.795	6.44	8.165	ERR	78.85
Ph IV	1.228	1.43	0.353	1.318	0.07	1.543	22.61	27.1	7	8.2	ERR	13	81.5	2.86	15.93	ERR	4.403	ERR	ERR	ERR	75.75
Ph V	0.541	0.616	0.234	0.92	ERR	2.386	15.08	23.33	3.5	3.375	ERR	19.31	75.5	3.29	28.03	3.949	4.88	7.113	9.165	18.57	77.29

After May 15, all three brush aerators were put in service. The level of liquid in the oxidation ditch was increased to maximize the depth of submergence and the amount of air transferred by each brush aerator. Nitrification increased over a two week period and the plant achieved complete conversion of ammonia to nitrates by the end of May.

The lack of oxygen resulted in an increase in effluent BOD_5, SCOD, COD, unoxidized nitrogen forms and phosphorus. The results are shown in Figures 6, 7, 8 and 9. The phosphorus levels increased because ferrous iron was not oxidized completely to ferric form. The ferrous iron was oxidized in the post aeration basin where it formed a brown precipitate. Several surfactants escaped breakdown and treatment and formed copious quantities of foam in the post aeration basin. The quality of the effluent improved after the supply of oxygen was increased to satisfy the demand (at the end of May, 1988). The problem with the foam was not eliminated altogether until the process removals increased in Phase IV.

For a few days at the end of July, the aeration was increased to examine the possibility of alkalinity limitation on nitrification. Complete nitrification was maintained at alkalinity levels of 35 mg/L and pH levels of 6.4 at an MCRT of 10 days. The liquid temperature was 21 Celsius.

The SVI increased from 125 mL/g in mid-June, 1988, to 200 mL/g at the end of July. The increase was associated with an increase in the population of certain types of filamentous bacteria (Sphaerotilus Natans). SVIs increased when soluble COD levels increased due to the failure of one or more aeration devices over 12 to 36 hour periods. The growth of filamentous bacteria is discussed later. A return sludge chlorination line was not available to control the growth of filamentous bacteria.

Phase III (August 1988 to January 1989)

The plant was operated with one oxidation ditch during this Phase. The potential for biological phosphorus removal was examined after chemical phosphorus removal was discontinued since August 17, 1988. The results are

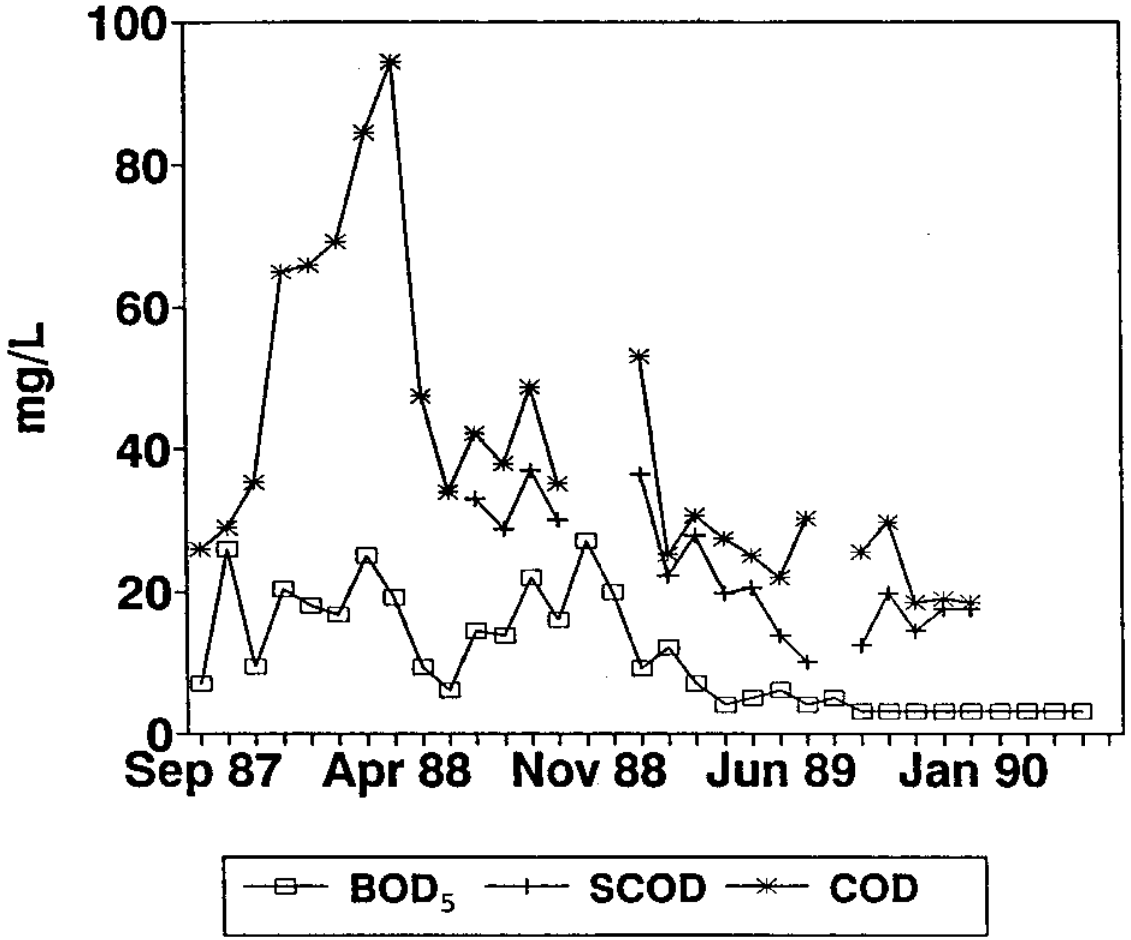


Figure 6. BOD_5, Soluble COD and Total COD Trend, Phase I to Phase V

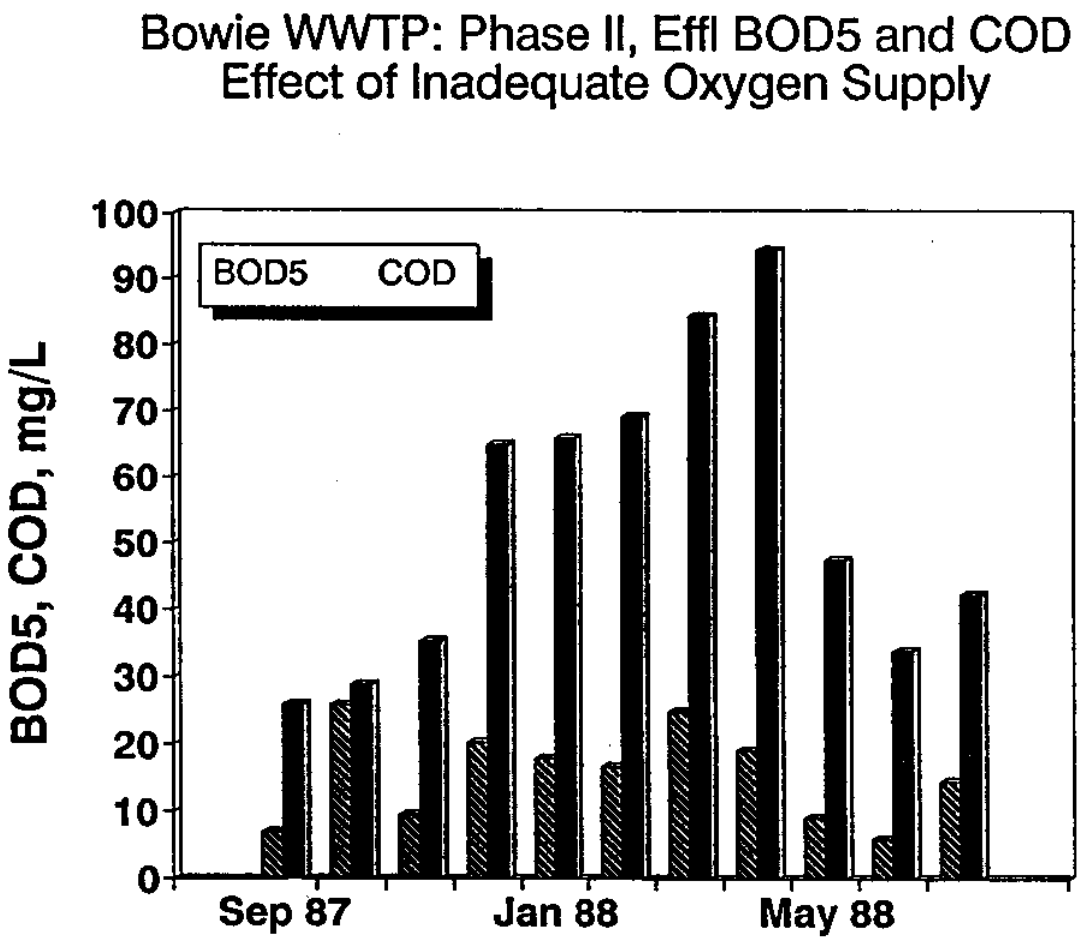


Figure 7. Effluent BOD_5 and COD in Phase II

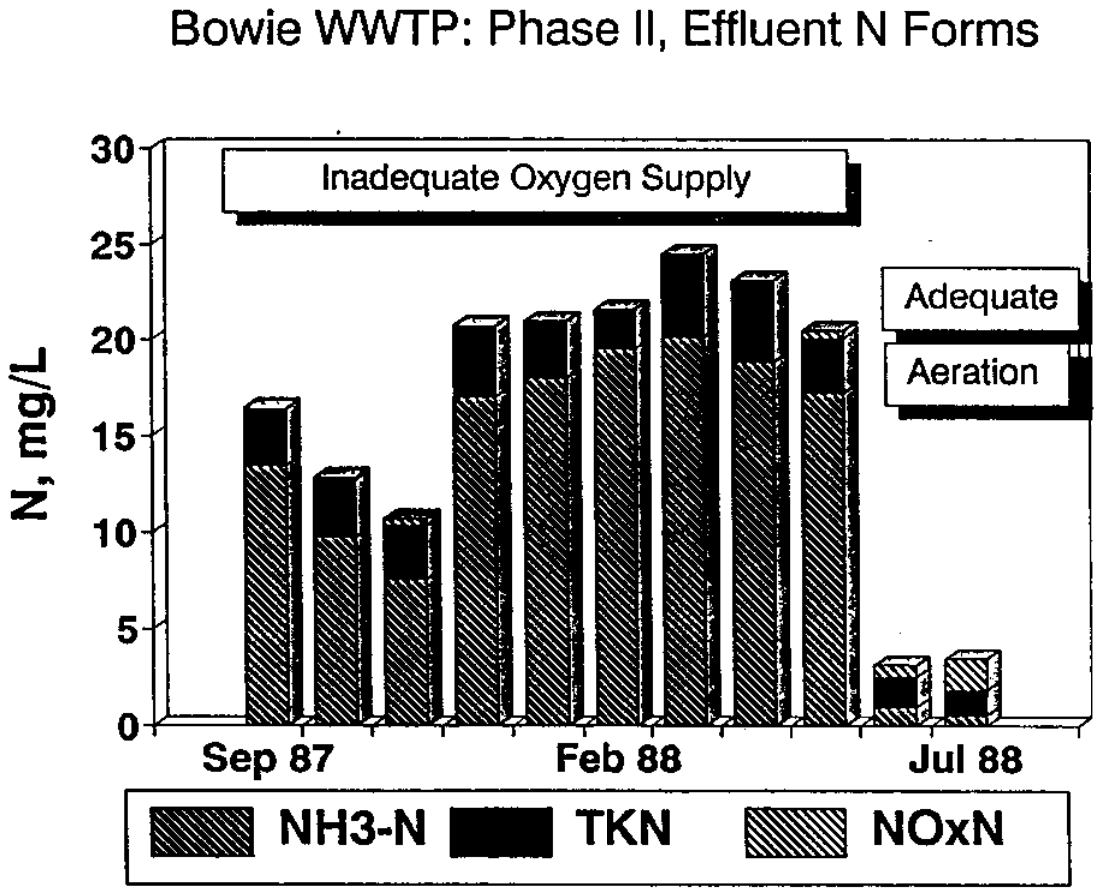


Figure 8. Effluent Nitrogen Forms in Phase II

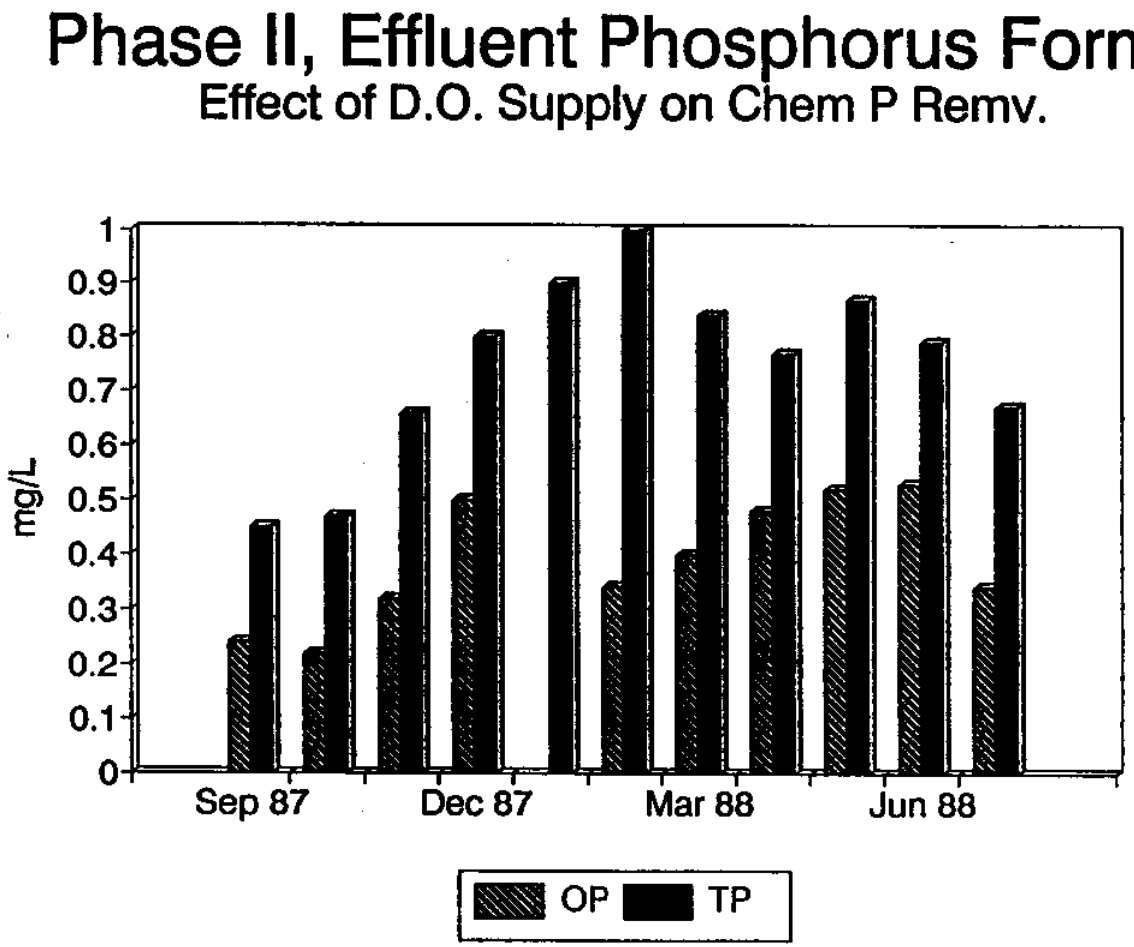


Figure 9. Effluent Phosphorus Forms in Phase II

presented in Tables 4 to 7. The Phase averages for phosphorus are computed for the results observed between September 1, 1988 to January 10, 1989.

Effluent total phosphorus increased to 0.67 mg/L in July to 2.2 mg/L in September. However, biological P removal increased gradually from September to November 1988 as with better control over the distribution of aerobic, anoxic and anaerobic volumes in the oxidation ditch. The improvement in effluent phosphorus is shown in Figure 10 and Table 7.

With chemical P removal discontinued, the alkalinity consumption decreased by 25 mg/L. A temporary return sludge chlorination line was installed in September 1988. Chlorination at the rate of 4 pounds per thousand pounds of volatile solids, an increase in operating pH, and control over anoxic and aerobic volumes resulted in a gradual reduction in the population of filamentous bacteria and an improvement in the SVI from 220 mL/g in August to 160 mL/g in September (Table 4).

In September, 1988, one of the brush aerators was taken out of service for repairs. The ammonia levels increased because the other aerators could not transfer an adequate amount of oxygen to satisfy the demand (Figure 11). Effluent ammonia-N levels improved after the aerator was put back in service. This verified that if the plant were operated with one oxidation ditch only, it would be difficult to satisfy ammonia and total nitrogen permits unless standby aeration devices were available. Installation of standby aeration devices would require new electrical and mechanical equipment and were outside the scope of this project.

The effluent ammonia levels increased from 0.18 mg/L in November to 5.84 mg/L in December, 1988 (Figure 11 and Table 7). Two changes were made in an attempt to increase nitrification. The MLSS in the oxidation ditch was increased from 3600 mg/L to 4100 mg/L, increasing the MCRT from 12 to 14 days in the process. Since nitrification is controlled by the aerobic MCRT, the amount of air injected was also increased to increase the aerobic volume. Though this helped increase nitrification from December to January, the level of nitrates increased because the anoxic volume (and MCRT) was not adequate. The clarifier solids loading exceeded 25 pounds per

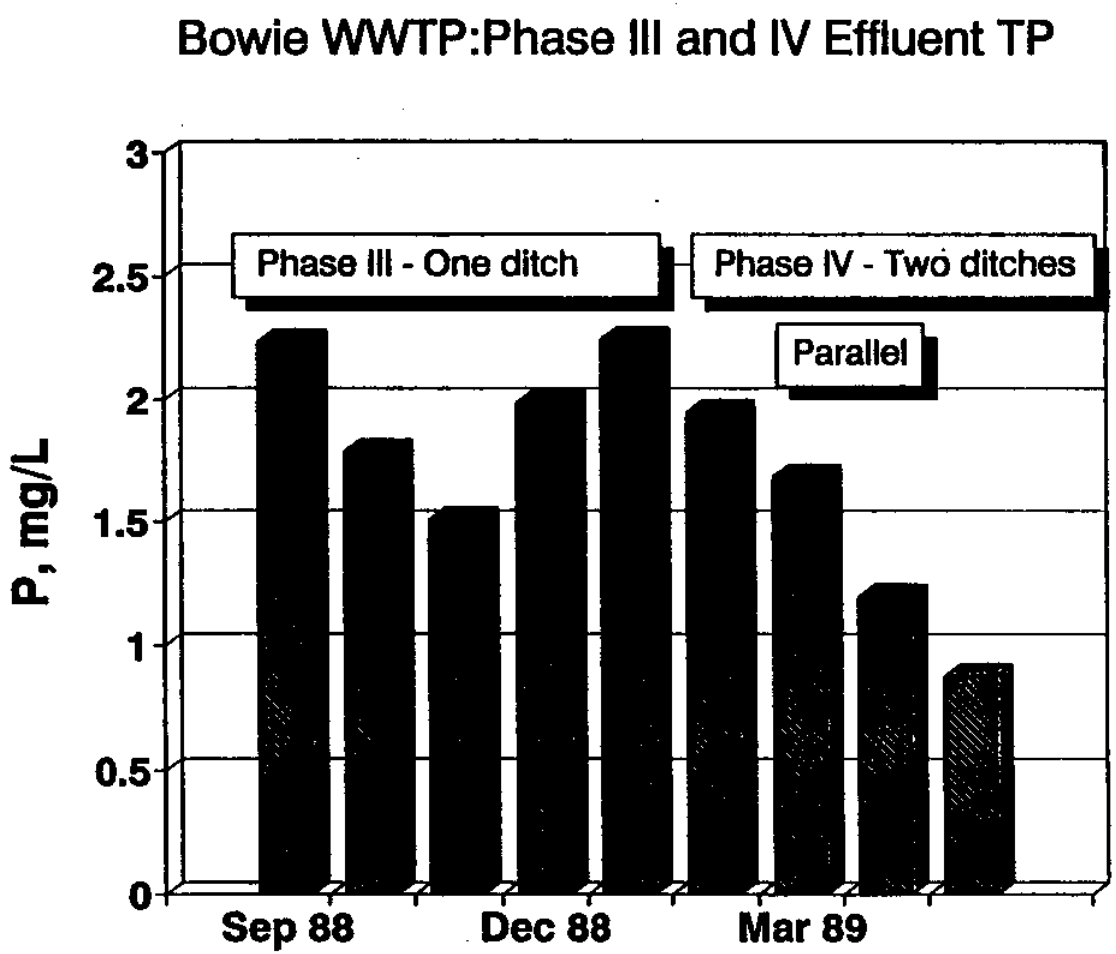


Figure 10. Effluent Phosphorus Forms in Phases III and IV

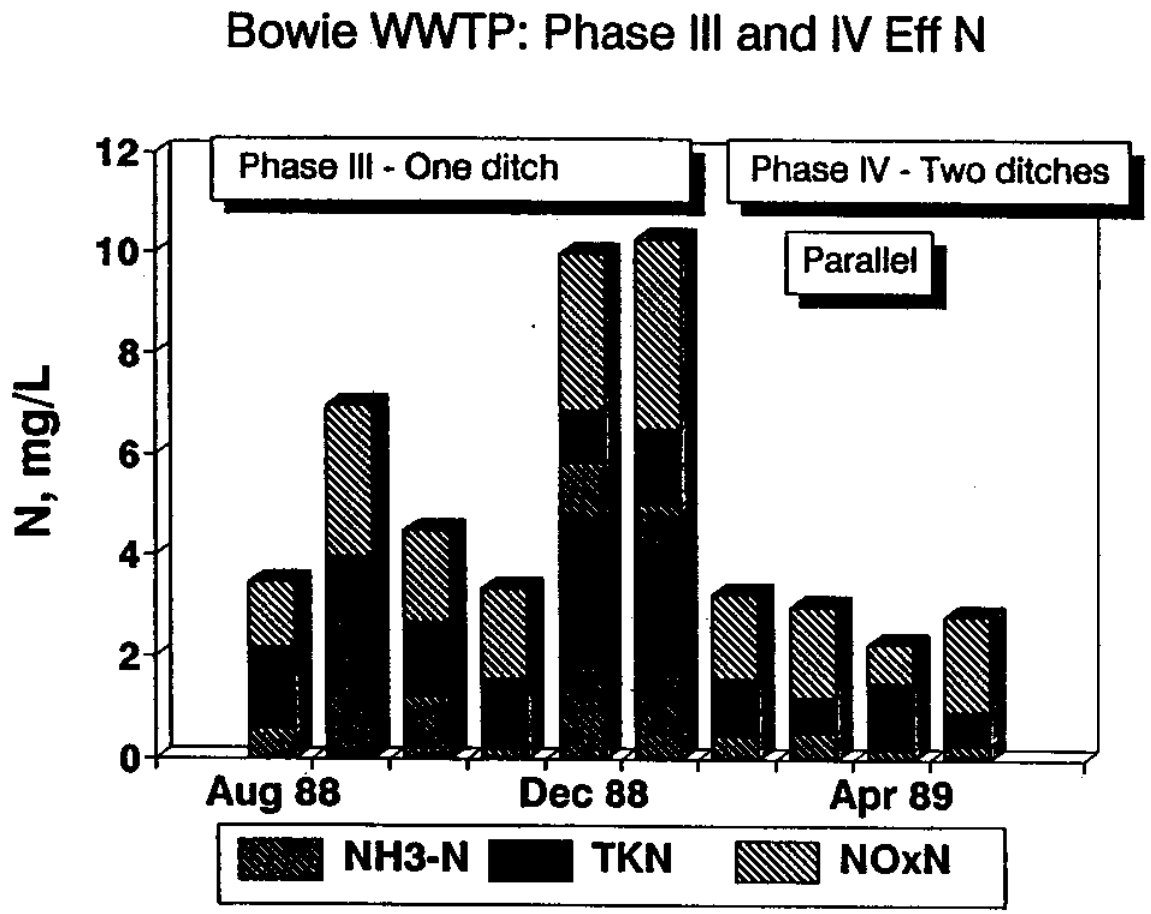


Figure 11. Effluent Nitrogen Forms in Phases III and IV

square foot of surface area between 0800 to 2200 hours each day. An increase in the effluent suspended solids at the high clarifier loadings increased the organic-N concentrations in the effluent.

In January, 1989, it was evident that complete nitrification with good denitrification would require operation with two oxidation ditches. Much of the digested sludge in the second oxidation ditch was dewatered over a 30 day period between January 15 and February 15, 1989 to make the second ditch available for treatment puposes.

Phase IV (February 1989 to May 1989)

During this phase, the plant was operated with the two oxidation ditches operating in parallel. In April, 1989, one half of the flow equalization basin was modified to an anaerobic cell and connected to oxidation ditch 1 (Figure 12). However, the piping from the anaerobic cell was not extended to oxidation ditch 2. The results from Phase IV are presented in Tables 4 to 7.

When flow was introduced into the second oxidation ditch, there was an immediate reduction in the effluent ammonia-N from 5 mg/L to less than 0.5 mg/L. Since oxidation ditch 2 had been used as an aerobic digester, it contained a large population of nitrifiers. When some of the digested sludge was mixed with the biomass in oxidation ditch 1, the population of nitrifiers was sufficient to convert all the ammonia to nitrate.

This mode of operation reached "steady state" in March, 1989. The system was operated at an MCRT of 15 to 20 days. During this phase, it became evident that it would be extremely difficult to control the aeration in the oxidation ditch system using D.O. as the sole control parameter. This was because D.O. measured less than 0.5 mg/L at all accessible points in the oxidation ditch. Two alternative parameters were evaluated. The oxidation ditch and secondary effluent alkalinities were used as the primary control parameter, and oxidation ditch supernatant and secondary effluent turbidities were used as a secondary control parameter. These parameters

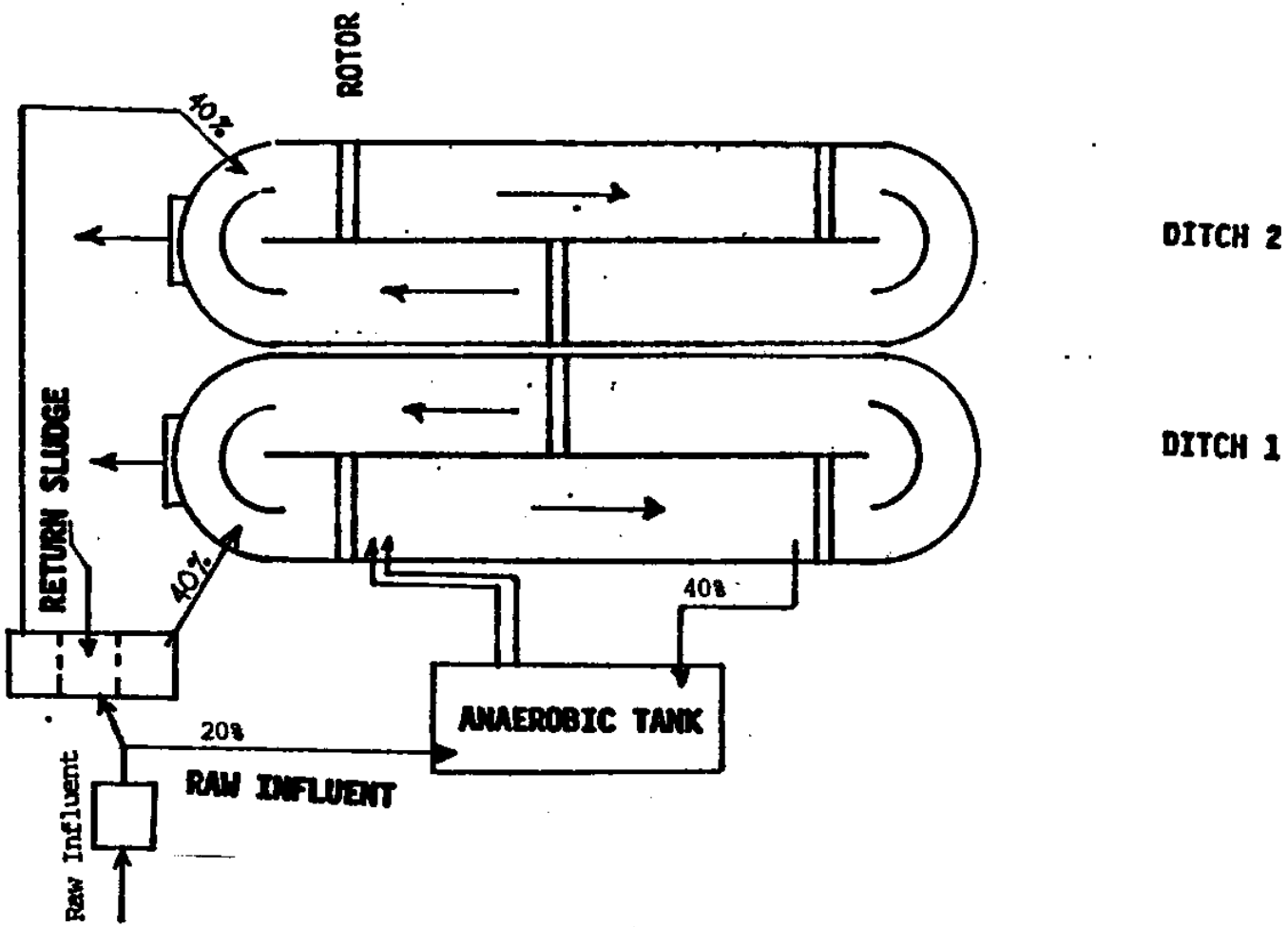


Figure 12. Flow Schematic for Phase IV

and their equivalence to the amount of oxygen supplied are discussed in the following chapter.

There was some improvement in the effluent phosphorus levels when both oxidation ditches were operated in parallel. The effluent phosphorus improved from 1.7 mg/L in March 1989 to 0.9 mg/L in May (Table 7 and Figure 10). The effluent soluble COD decreased from an average of 35 mg/L in Phase III to less than 25 mg/L in Phase IV (Figure 6).

For the flow arrangement utilized in Phase IV, the anaerobic cell was connected to oxidation ditch 1 only (Figure 12). As a result, the raw influent was not distributed equally between the two oxidation ditches. If 20 percent of the influent flow was diverted to the anaerobic cell at the grit chamber and then pumped to oxidation ditch 1, the other 80 percent was distributed equally at the flow splitter box between the two ditches. The return sludge was also distributed equally between the two ditches. This resulted in a higher solids concentration in ditch 2 than in ditch 1. 60 percent of the influent flow was allowed to mix with 50 percent of the return sludge in ditch 1 while only 40 percent of the influent flow mixed with the other 50 percent of the return sludge in ditch 2 (Table 8). Calculations presented in Table 8 show that the resulting difference in the amount of biomass generated during growth compensated for only 5 percent of the difference in MLSS at a nominal HRT of 36 hours and an operating MCRT of 20 days.

Phase V (June 1989 to May 1990)

In May, 1989, final approval was obtained from the Maryland Department of Environment to connect the two oxidation ditches in series. A 10 foot wide notch was drilled below the minimum level of liquid which could be maintained in the oxidation ditches. Construction was completed during the first two weeks of June and the ditches were connected in series (Figure 1).

Table 8. Solids Distribution Between Ditches in Phase IV

	Ditch 1	Ditch 2
1. Influent Flow	60%	40%
2. Influent Flow	1.8 MGD	1.2 MGD
3. Return Solids	50%	50%
4. Return Flow	1.0 MGD	1.0 MGD
5. Return MLSS (mg/L)	7500	7500
6. Actual HRT	14.14 hours	18 hours
7. Ditch MLSS from Return sludge[1]	2680 mg/L	3410 mg/L
8. Biomass generated[2]	115 mg/L	98 mg/L
9. Total MLSS	2795 mg/L	3508 mg/L
10. Loss due to decay[3]	82 mg/L	132 mg/L
11. Biomass from decay[4]	54 mg/L	84 mg/L
12. Aeration Effluent Biomass	2767 mg/L	3460 mg/L
13. Second Iteration	2766 mg/L	3461 mg/L

[1]MLSS from return sludge = $\frac{\text{Return Sludge MLSS} * \text{(Return flow to ditch)}}{\text{(Return Flow + Influent Flow to ditch)}}$

[2]mg/L of biomass produced = $\frac{\text{Influent Flow} * 400 \text{ mg/L COD} * 0.45 \text{ (Yield)}}{\text{(Return Flow + Influent Flow to ditch)}}$

[3]Decay of biomass (mg/L) = MLSS * Decay rate/day * actual HRT (day)

[4]Biomass from COD released by decay (mg/L) = Decay biomass (mg/L) * 1.42 * 0.45

Second Iteration (13) calculates new values for decay (10) and the biomass from decay (11) using the aeration effluent biomass from the first iteration (12) as the Total MLSS (9).

The operating procedures were developed after analysis of the performance in Phases II, III and IV. The Operating Strategy is outlined below and is summarized in Table 9.

General Operating Principles

1) Sufficient oxygen has to be transferred to complete nitrification in a biological system whose aerobic MCRT is adequate for nitrification at the liquid temperature of concern, i.e. the mixed liquor temperature.

2) The anoxic MCRT should be adequate for denitrification to achieve acceptable total nitrogen levels at the liquid temperature of concern.

3) To the extent that it is feasible, the aeration in ditch 2 (to complete BOD stabilization and nitrification, with simultaneous credit for denitrification) should be maximized. The aeration in ditch 1 should be determined by the mixing requirements.

4) If the air supplied in ditch 2 is at its maximum available capacity, the difference between the requirement (for BOD stabilization and nitrification with credit for denitrification) and that supplied in ditch 2 should be supplied in ditch 1. The maximum aeration capacity should determined after allowing for sufficient standby aeration capacity (0.5 to 1 brush aerator under normal circumstances).

5) Every attempt should be made to maintain anoxic or anaerobic conditions in the first leg of ditch 1. Unless brush aerators 1-2 and 1-3 are both out of service, it is recommended that aerator 1-1 not be operated at all (Figure 1). Spare gear boxes, shafts and bearings should minimize the probability of simultaneous failure of two aerators.

6) Unless maintenance activities dictate, aerator 1-2 should be operated all the time. Aerator 1-3 should be operated continuously when aerator 1-2 is out of service.

Table 9. Aeration Strategy, Concepts and Instructions

Air Requirement = Requirement for BOD stabilization
+ Requirement for nitrification
- Credit for denitrification.

Air Supplied = Air Supplied in ditch 2 + Air Supplied in ditch 1.

Conditions: 1) a) Maximize air supplied in ditch 2 while maintaining denitrification

b) Minimize air supplied to ditch 1 but ensure adequate mixing

2) Operate at least two aerators in ditch 2 and one aerator in ditch 1 all the time.

3) Flow pace aeration using one aerator in ditch 1 and/or one aerator in ditch 2.

4) Maintain unaerated section in the first pass in ditch 1. (Figure 1).

Down time situations:

One aerator in ditch 1 is out of service -

1) Maximize aeration in ditch 2 by increasing liquid level in ditch and operating two aerators all the time. The off time on the third aerator should be minimized (0 to 6 h).

2) If aerator 1-2 is out of service, aerator 1-3 should be operated all the time. Do not flow pace in ditch 1.

3) Ensure adequate aeration for weekend high flows. Raise weir or increase the on time on third aerator in ditch 2 by 4 h. up to a max. of 24 h. during weekends.

4) Aerator 1-1 should not be operated unless 1-2 and 1-3 are both out of service. P removal will deteriorate and filamentous bacteria will increase if 1-1 is operated.

Aerator in ditch 2 out of service

1) Raise weir as far as possible and operate two aerators in ditch 2 all the time.

2) Bring alkalinity to operating range by operating one aerator ditch 1 in ditch 1 all the time (1-2) and use a second aerator (1-3) to flow pace aeration.

Simultaneous failure of one aerator in each ditch:

Operate aerator 1-1 part of the day as a last resort.

7) At least two aerators in ditch 2 and one aerator in ditch 1 should always be in operation. A third aerator in ditch 2 and/or a second aerator (aerator 1-3) in ditch 1 should be used to flow pace aeration.

8) The level of liquid in ditch 2 should be adjusted up to a maximum of three times each day to ensure that the air transferred by the aerators (in both ditches) is within a range which is adequate for nitrification and denitrification. Alkalinity and turbidity measurements on the supernatant from ditch 2 and the secondary clarifier effluent should be used as the control parameters to make this determination. Alkalinity and turbidity measurements on the secondary effluent should be conducted two times each day. The alkalinity in ditch 2 should be measured once each day on the supernatant from the sludge settlometer test.

9) The aeration should be increased during weekends when the plant receives higher flows. Results from Monday afternoon should be used to determine the magnitude of the adjustment for the following week.

10) Aeration should be increased for holidays if higher flows are expected.

Operating Principles for Temperature Compensation

A key focus of this research was to determine operating procedures which would allow oxidation ditches (and activated sludge processes in general) to maintain performance levels in cold weather which would be similar to those observed at warmer temperatures. The following reactions are affected by temperature:

a) Rates of BOD storage and stabilization which affect specific oxygen uptake rate;
b) Rate of nitrification which affects specific oxygen uptake rate;
c) Rate of denitrification which affects specific nitrate uptake rate;
d) Rate of fermentation in sewers which will determine the amount of readily available BOD available for storage under anaerobic conditions; and

e) Oxygen transferred by each brush aerator which is affected by

i) the liquid temperature (determines the solubility of oxygen in the wastewater), and

ii) atmospheric temperature (determines the partial pressure of water vapor and air in the saturated layer of air above the water surface).

The following adjustments have to be made to compensate for temperature changes:

1) With a seasonal drop in temperature, the MLSS concentrations are increased to increase the oxygen and nitrate uptake rates (mg/L-hr), thereby compensating (partially) for the drop in specific oxygen and nitrate uptake rates.

2) Dissolved oxygen levels in the grit chamber are minimized to reduce the amount of oxygen in the influent to the anaerobic zone.

3) Because of reaeration of liquid in the anaerobic cell at the points where the influent from the grit chamber and the mixed liquor recycle from ditch 1 enter the anaerobic cell, these flows are reduced in winter. This increases the anaerobic detention time and compensates for the drop in oxygen uptake, nitrate uptake and fermentation rates.

4) A greater fraction of the air is supplied in ditch 2. The operating hours for the aerators in ditch 1 are decreased and/or those in ditch 2 are increased while maintaining good nitrification and denitrification. This helps maintain sufficient pounds of biomass under anaerobic conditions in ditch 1 and sustain biological phosphorus removal.

Increasing the air supplied in ditch 2 reduces the anoxic volume. This may result in an increase in effluent nitrates from levels below 2 mg/L in summer to 3 mg/L in winter. The system MCRT is increased if nitrate levels exceed 3 mg/L to increase the anoxic MCRT and denitrification in ditch 2.

Temperature determines the operating MCRT range for any activated sludge system because it controls

a) the minimum MCRT at which nitrification, denitrification and biological P removal can be sustained; and
b) the maximum MCRT beyond which the capacity of the aeration system is exceeded.

It is preferable that the "operating window" between the minimum and the maximum MCRTs be large enough to avoid process limitations from sudden changes in temperature. At a flow of 2.2 MGD, the Bowie WWTP can maintain MCRTs between 15 and 25 days in summer and up to 40 days in winter with one standby aerator. Since these are above the minimum MCRTs required for nitrogen and phosphorus removal, the plant can operate reliably at this flow. Unless aeration capacity is increased, the operating window will shrink at higher flows.

The results for a 12 month period of operation in the VT2 process mode are presented in Tables 4 to 7. Effluent phosphorus levels averaged 0.6 mg/L in cold weather and 0.5 mg/L in warm weather (Figure 13). The performance will improve if process control can be enhanced. Effluent phosphorus levels in 1990 are expected to average 0.1 mg/L less than those observed in 1989.

In September, 1989, the anaerobic cell was taken out of service. Phosphorus levels increased to 1.2 mg/L. The experiment was discontinued after September, 1989, because the plant was expected to meet a permit of 1 mg/L for total phosphorus. The SVI increased during this period (Figure 14).

During Phase V, the monthly average nitrogen levels were maintained below 0.5 mg/L for ammonia-N, at 1 mg/L for TKN and 3 mg/L for total nitrogen (Figure 15). The process is being optimized to maintain effluent total nitrogen levels below 3 mg/L in warm weather and 4 mg/L in cold weather.

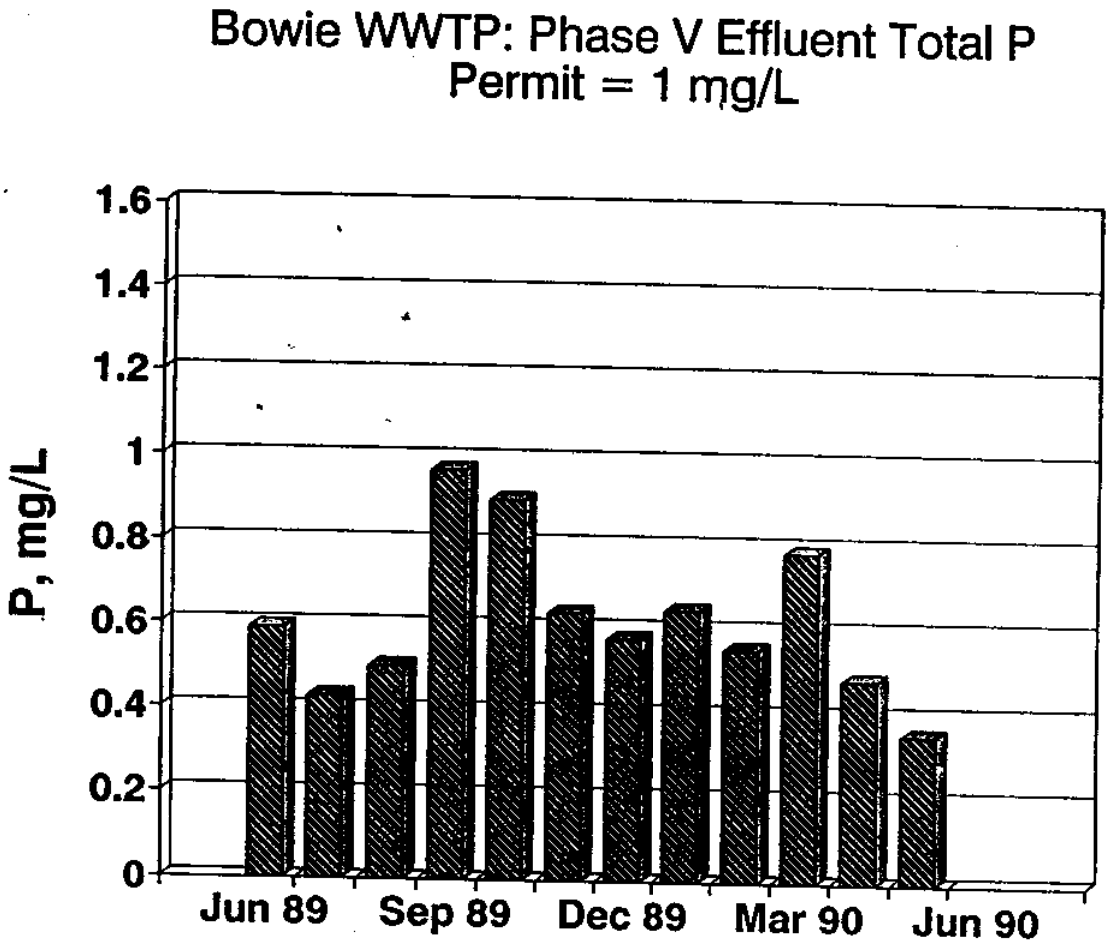


Figure 13. Effluent Total Phosphorus, Phase V

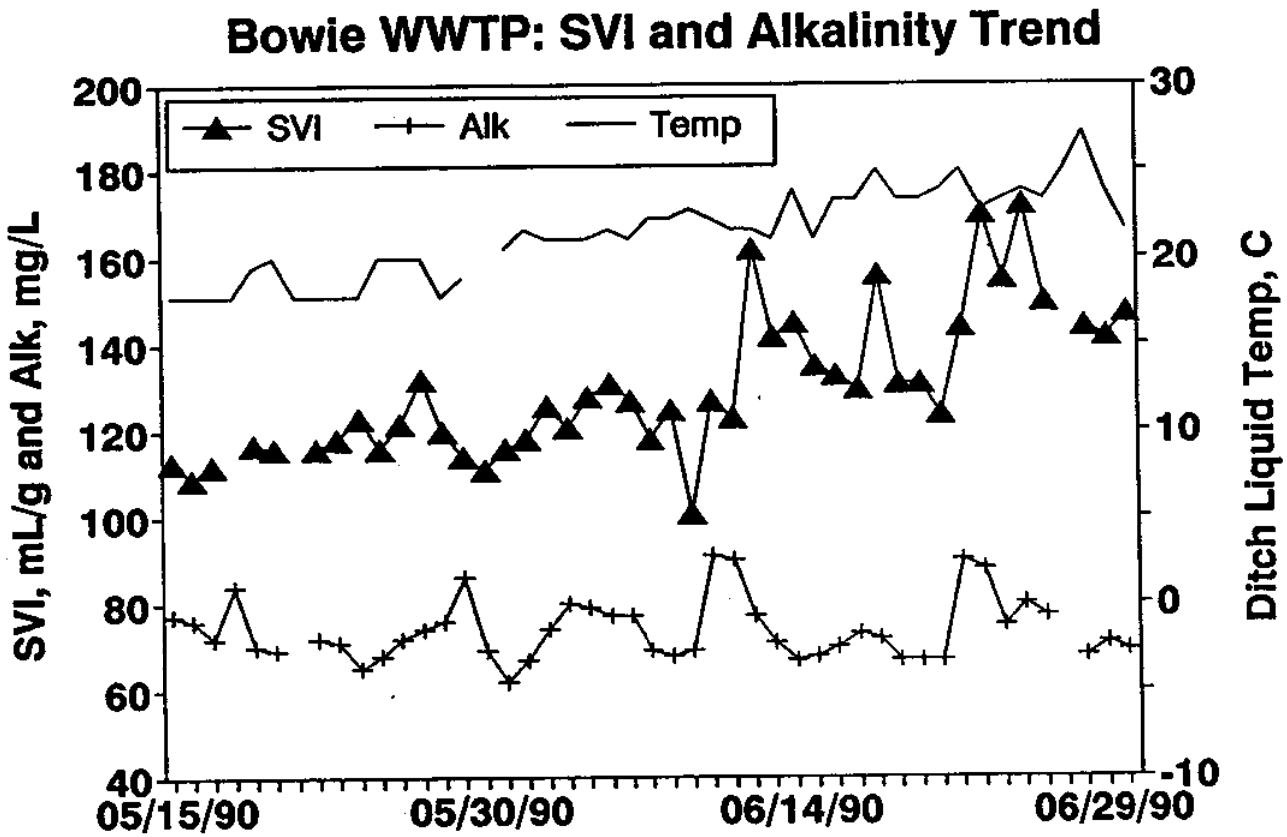


Figure 14. SVI and Alkalinity Trend, May-June 1990

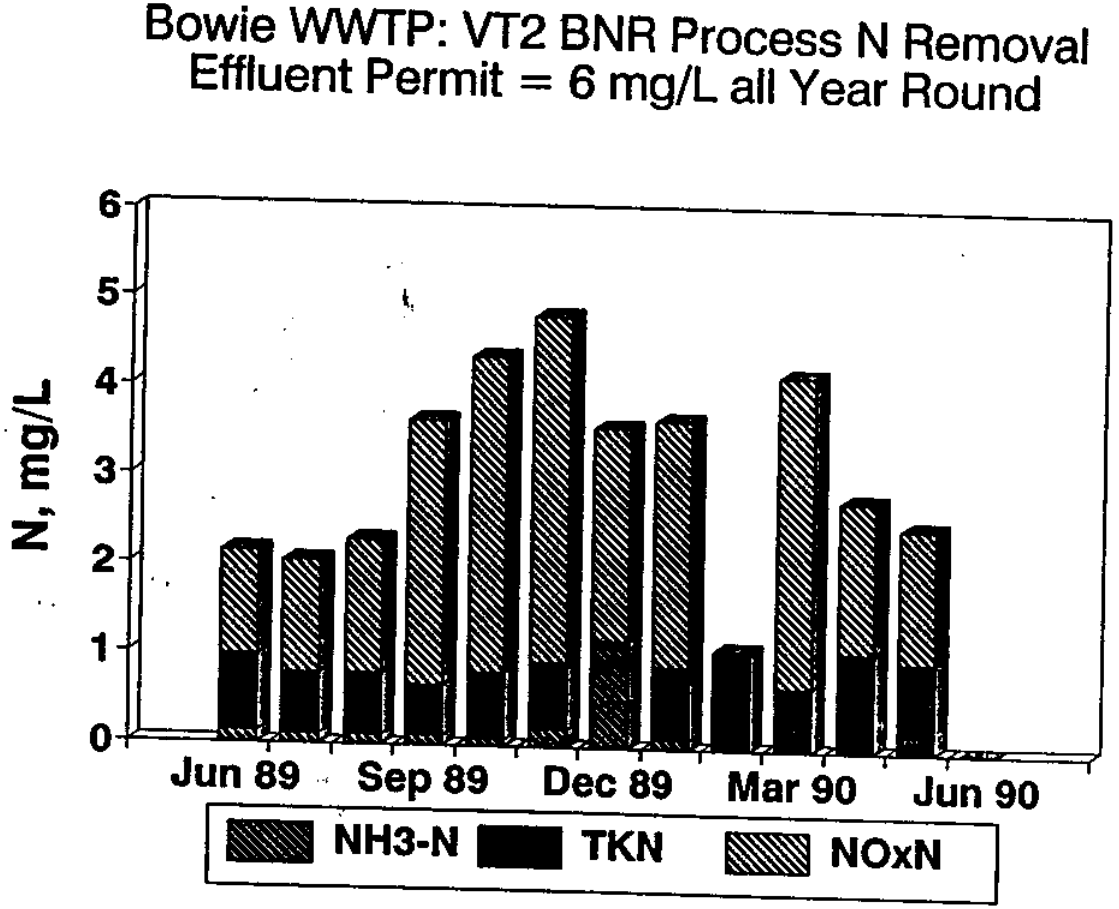


Figure 15. Effluent Nitrogen Forms, Phase V

Phase VI

Phase VI is designed to increase the efficiency of the VT2-BNR process. The following upgrades are planned:

1) Two chutes will be installed to enclose the liquid flowing through the influent and mixed liquor recycle lines at the points where it enters the anaerobic cell;

2) The pipes carrying liquid from the anaerobic cell to oxidation ditch 1 will be extended below the surface of the liquid to eliminate reaeration at the oxidation ditch;

3) A chlorination system will be installed to chlorinate the oxidation ditch at multiple points. This will allow the biomass to be exposed to chlorine at least three times each day if and when there is a problem with settling rates because of a high population of filamentous bacteria. The actual HRT of the biological system (oxidation ditches, clarifiers and anaerobic cell) is 22 hours when 4 MGD of flow (influent + return sludge flow). The biomass is exposed to chlorine only once each day if chlorine is injected into the return sludge stream. This frequency is not adequate for controlling the population of filamentous bacteria.

4) Increasing the level of the effluent weir in ditch 1 will eliminate the overflow across it. In the present configuration, mixed liquor from ditch 1 overflows the weir during high flows, short circuiting ditch 2 in the process, and releases ammonia and unstabilized BOD into the clarifiers.

5) The timers on each brush aerator will be improved.

It is expected that the phosphorus levels will decrease by 20 to 30 percent below the present day averages of 0.5 mg/L for summer and 0.6 mg/L for winter. Effluent nitrogen levels will be maintained at 3 mg/L total nitrogen in summer and 4 mg/L total nitrogen in winter. The system will be able to handle 2.5 MGD of flow at BOD_5 levels of 160 mg/L and COD levels of 400 mg/L using the available aeration and solids handling capacities.

DISCUSSION

The biological nutrient removal system developed at the Bowie WWTP was designed to optimize nutrient removal within the constraints of the basin structure and the location of the aeration devices. The biochemical reactions involved in BOD removal, nitrification, denitrification and excess biological phosphorus uptake are similar to those in other activated sludge systems. The following improvements in process performance were achieved by optimizing aeration, mixing and recycle rates:

a) reduced effluent ammonia-N from levels above 10 mg/L to less than 0.5 mg/L (monthly average) all year round;

b) reduced effluent TKN from levels above 10 mg/L to 1 mg/L (monthly average) all year round;

c) reduced effluent nitrates from levels above 15 mg/L, as would be observed in the absence of denitrification, to than less 2 mg/L all year round without the use of methanol;

d) eliminated the need for supplemental alkalinity (which would be necessary if the denitrification had not reduced effluent nitrate levels below 5 mg/L);

e) reduced effluent phosphorus levels from 3.0 mg/L to 0.5 mg/L using excess biological P removal with no chemical supplements; supplemental alkalinity requirements were cut back in the process; and

f) achieved better control over filamentous bacteria and settling rates by controlling anaerobic, anoxic and aerobic volumes within the oxidation ditches.

Process operation for the VT2-BNR process has been discussed earlier. Several guidelines related to its design and operation were developed after experimental work on BNR processes in other oxidation ditches and in modified activated sludge tanks.

Process Control Parameters

Dissolved Oxygen

Dissolved oxygen is the best known control parameter for aeration in activated sludge systems. Oxygen is essential for BOD stabilization and for the conversion of ammonia to nitrate. In a single stage system which has an adequate population of nitrifiers, the oxygen supplied is considered to be adequate if the ammonia level in the effluent is below the desired operating average. The minimum operating dissolved oxygen level (MODO), as measured at a particular location in the activated sludge basin, is the minimum set point at which satisfactory BOD stabilization and nitrification are achieved.

Several parameters will influence the MODO for nitrification. It should be noted that at the MODO, the aerobic MCRT is equal to the minimum required for nitrification. The minimum nitrifier MCRT is influenced by:

a) Operating D.O. levels - when nitrifiers grow at D.O. levels below that required to achieve their maximum growth rate (u_m), the minimum nitrifier MCRT will have to be increased to compensate for the reduction in growth rates (u);

$$u = u_m \left[D.O./(K_{D.O.} + D.O.) \right] \left[NH_4\text{-}N / K_N + NH_4\text{-}N \right]$$

$$\theta_{aerobic} = 1/u$$

where

D.O. = operating D.O. level for a completely mixed system;

$K_{D.O.}$ = half saturation constant for dissolved oxygen on nitrification;

NH_4-N = ammonia-N concentration

K_N = half saturation constant for ammonia-N.

$\theta_{aerobic}$ = MCRT required for nitrification

These equations are valid for a completely mixed system. They have to be modified for an oxidation ditch in which the D.O. decreases with distance from the aerator. The ammonia-N level remains fairly constant because of high internal dilutions. Though the $K_{D.O.}$ may be higher in sections of the oxidation ditch where higher concentrations of unstabilized BOD are available (Hanaki et.al., 1990 a,b), the computation may be performed for a constant $K_{D.O.}$ because high internal recycle rates result in fairly uniform BOD levels across the oxidation ditch. A computer model can be developed to simulate nitrification across thin sections of the oxidation ditch. A plug flow equation can be developed if $K_{D.O.}$ and K_N are held constant. Appropriate equations will be developed in the near future.

b) Operating temperature - at lower liquid temperatures, the minimum nitrifier MCRT will have to be increased to compensate for the reduction in the growth rate of the nitrifiers;

c) Operating ammonia-N concentration after dilution with recycle flows containing low levels of ammonia-N;

This may occur in oxidation ditches where the ammonia-N concentration is diluted by the internal recycle (flow through a cross section of the ditch) to levels below 1 mg/L immediately after the influent is introduced into the oxidation ditch (0.04 to 0.5 mg/L NH_4-N). Low levels of ammonia-N may reduce growth rates and increase the minimum nitrifier MCRT.

d) Operating BOD concentration after dilution of influent BOD with the internal recycle;

This occurs in oxidation ditches where the BOD is diluted to levels close to those in the secondary effluent. As a result, nitrifiers face less competition for dissolved oxygen from heterotrophs in the first part of the aerobic zone. Nitrification may occur simultaneously with BOD stabilization. This may be represented mathematically as a reduction in $K_{D.O.}$ for nitrifiers and may help reduce the minimum nitrifier MCRT.

There is strong evidence that K_N values are lower in oxidation ditches because the nitrifiers face less competition for ammonia-N from heterotrophs. This can reduce the minimum MCRT required for nitrification and can offset the effect of (c) above. Low levels of ammonia-N can reduce the growth rate of heterotrophs in biological systems where the organic compounds are hydrolyzed prior to heterotrophic growth. Hydrolysis of TKN to ammonia-N prior to biological uptake may be prevail in BNR systems because of fermentation and enzymatic activity on substrates in the anaerobic cell(s).

The results from the oxidation ditches at Bowie and Patuxent have shown that nitrification and BOD removal can proceed simultaneously if dissolved oxygen is available. This indicates there is a reduction in $K_{D.O.}$ and K_N levels because of the dilution effect of high internal recycle rates, very similar to recent observations in CSTRs where the $K_{D.O.}$ was observed to be a function of influent BOD to TKN ratios (Hanaki et.al., 1990 a,b). Pilot scale research has to be conducted to quantify the reduction.

The of minimum operating D.O. level (MODO) for BOD stabilization, nitrification and denitrification is a function of several parameters.

1) For a basin with a constant volume, loading, flow and total system MCRT,

the MODO at a certain temperature is negatively correlated to the magnitude of return sludge recycle and nitrate recycle. These recycles increase the flow through a cross-section of the tank. For the same pounds of oxygen supplied per day, this dilutes the oxygen supplied per unit time into a higher volume of liquid. The actual hydraulic retention time is equal to the circulation time - the time required for the liquid to flow once around an oxidation ditch.

Circulation time = Volume of tank / (Influent + Return Sludge + Internal or Nitrate Recycle Flow Rates)

2) For a constant volume, loading, flow and total MCRT and recycle rates, the MODO will be a function of temperature.

If the MCRT of the biomass contained in the aerobic tanks is higher than the minimum required for nitrification, MODO will depend on existence of a D.O. gradient between the points of aeration.

If a D.O. gradient exists between the aerators, dissolved oxygen levels measured close to the aerator will increase with temperature. Because of high oxygen uptake rates, they will drop rapidly with distance from the aerators. The oxygen demand and oxygen uptake rates will decrease in winter if the system is operated at the same MCRT. As a result, D.O. levels measured close to the aerators may be less than those measured in summer. However, D.O. levels measured further away from the aerator may increase in winter (Figure 16). Therefore, the MODO will depend on the location of the D.O. probe relative to the point of aeration.

In a system which achieves complete mixing within the aeration tank (thereby eliminating any D.O. gradient between aerators), the MODO may decrease with an increase in temperature. Since nitrification rates increase with temperature, it may be possible to obtain complete nitrification under circumstances where the growth rate for nitrifiers

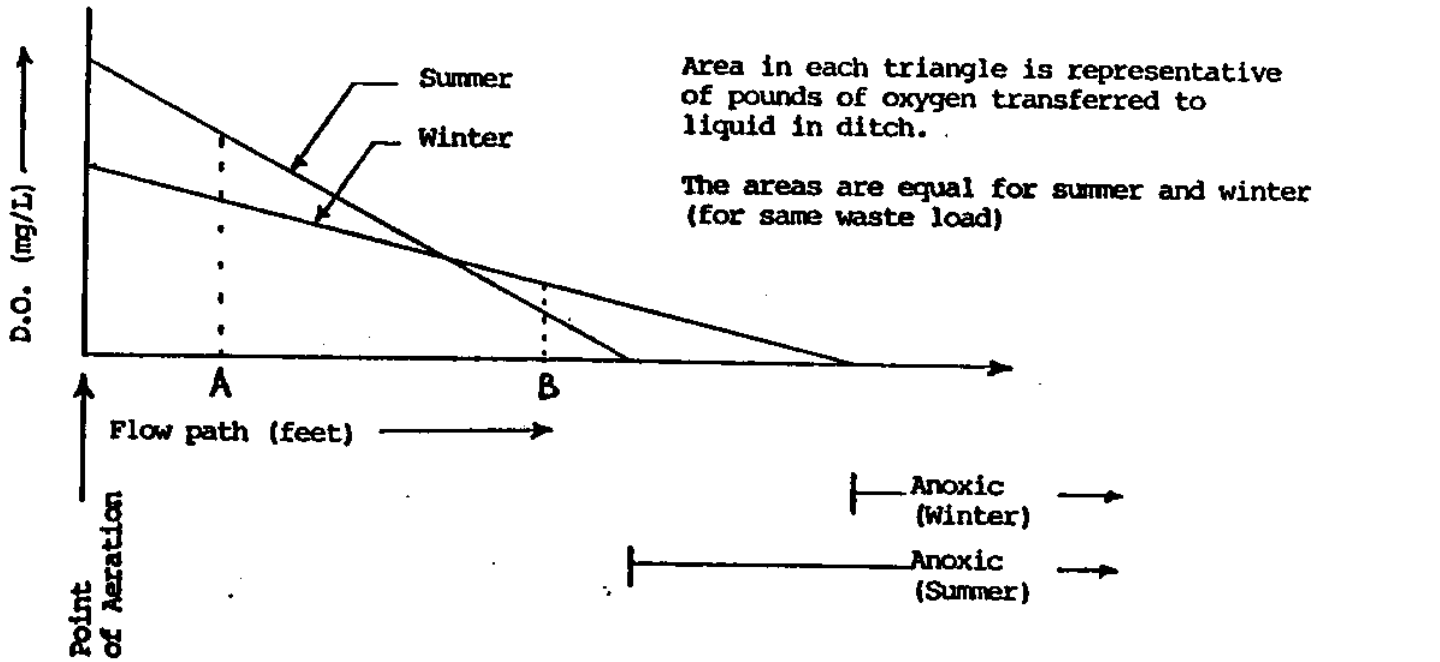


Figure 16. Effect of Temperature on Dissolved Oxygen gradient and the concentration of oxygen measured at different points along the flow path (under conditions of adequate oxygen supply).

is partially constrained by the dissolved oxygen concentration. In winter, as temperature reduces the rate of nitrification, dissolved oxygen levels cannot be used to constrain nitrification rates to the same extent as in summer. The MODO will then have to be increased in winter.

The aerobic MCRT will have to be increased by increasing the pounds of biomass under aeration if the MODO is well above the $K_{D.O.}$ for nitrifiers. This can be accomplished by increasing the MLSS or by introducing fixed film media in the aeration tank to increase the population of nitrifiers by cultivating them as attached growth.

3) For a system with constant loading, flow, MCRT and recycle rates, the MODO is a function of the volume of the aeration tank - i.e., the nominal HRT of the system. This is because the pounds of oxygen required have to be supplied within a smaller volume.

4) For a system with constant volume, loading, flow and temperature, the MODO will be a function of MCRT. The ratio of heterotrophs to nitrifiers is higher in a high rate system in comparison to a low rate system. Therefore, a high rate system (lower MCRT) may have a higher $K_{D.O.}$ for nitrifiers than a low rate system because of greater competition for oxygen from the heterotrophs. Soluble COD levels tend to be higher at low MCRTs. This can also increase $K_{D.O.}$ because it increases heterotrophic growth rates (Hanaki et.al., 1990a). Therefore, it is expected that the MODO will increase at a lower MCRT.

It is difficult to predict the change in the MODO when MCRT is changed in conjunction with temperature in a high rate system. Initial analysis indicates that the MODO may remain fairly constant all year round. Further research has to be conducted into this phenomenon.

A diffuse versus compact floc structure can alter the minimum set point. In a compact floc, the D.O. levels drop rapidly along the radius of the floc. However, in a diffuse floc structure, the gradient within the floc is not as high. When a system is operated at lower D.O. levels, floc structure

will be more diffuse. Filamentous bacteria tend to increase the diffuse nature of the floc (Jenkins et.al., 1986). It is possible that at operating levels close to the MODO, the population of filamentous bacteria may increase in the system. (Data will be presented later in the discussion which will relate the population of filamentous bacteria to the variation in control volumes under aerobic, anoxic and anaerobic conditions). Under these circumstances, the capacity of the clarifiers to settle sludge which contains a certain population of filamentous bacteria may have to be taken into account. A higher D.O. set point may be selected within the acceptable operating range to prevent the bacterial floc from attaining a diffuse structure.

5) In an oxidation ditch, the use of multiple points of aeration will reduce the MODO when anoxic conditions have to be created between the aerators to maintain denitrification. As the number of points for aeration are increased, it is possible that nitrification and BOD stabilization rates will be limited by lower dissolved oxygen levels existing over a larger fraction of the oxidation ditch. This will increase the minimum aerobic MCRT required for nitrification. Therefore, the minimum set point may not decrease linearly with the number of points of aeration. If the minimum set point measured adjacent to a point of aeration (y-axis) is plotted against the number of points of aeration (x-axis), its slope will be asymptotic to the x-axis (Figure 17). The maximum number of points of aeration will be limited by the minimum amount of biomass which must be maintained under anoxic and anaerobic conditions within the volume available.

The situation is complex and attempts have not been made to quantitatively predict the effect of a combination of factors on the MODO in oxidation ditches or in completely mixed aeration tanks. At most facilities, D.O. probes for process control are located close to the aerators. Under these conditions, the minimum operating D.O. level will increase with temperature, provided that complete nitrification is maintained all year round. This has been observed at the Patuxent Water Reclamation facility (barrier oxidation ditch) and the York River WWTP (UCT/VIP and A^2/O process). However, at Bowie WWTP and at Seneca WWTP in Maryland, where

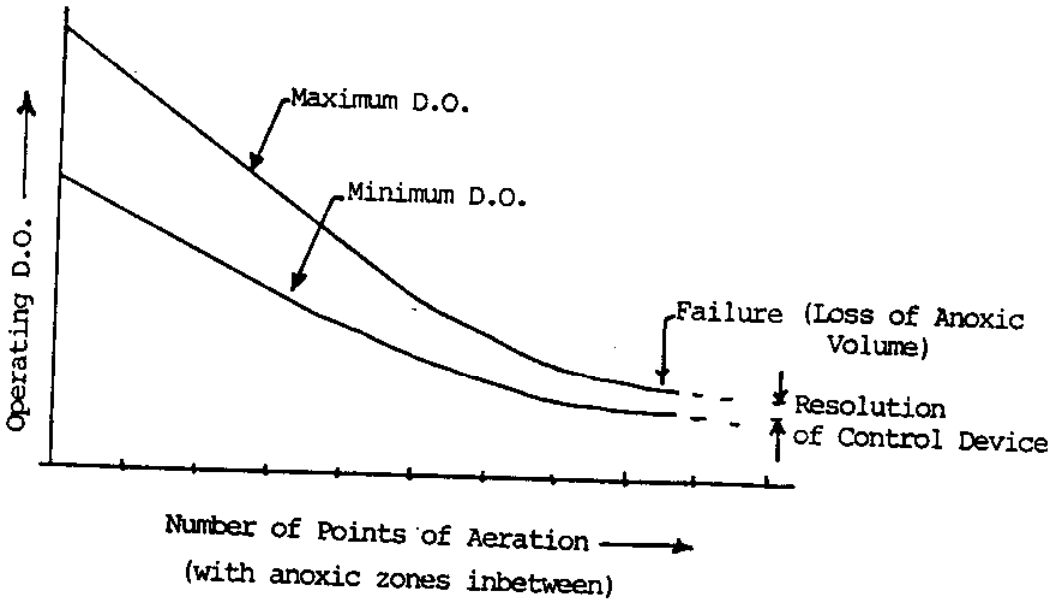


Figure 17. Minimum and Maximum Operating D.O. plotted against Number of Points of Aeration

anoxic conditions are established between points of aeration, the reduction in the D.O. gradient with the drop in temperature increased MODO levels in winter because measurements are taken further away from the aerator (Figure 16).

Since the MODO drops and the magnitude of the operating range shrinks in size as the number of points of aeration is increased, it becomes increasingly difficult to automate aeration and flow pace it off dissolved oxygen measurements (Figure 17). Such a situation has been experienced at Bowie, where, with as many as five points of aeration between the two ditches, D.O. measures less than 0.5 mg/L at accessible points of measurement (Figure 18 and Table 6). Therefore, alternative process control parameters which would be equivalent to the amount of oxygen supplied were evaluated.

Alkalinity as a Control Parameter

Model

During Phase IV, the alkalinity of the oxidation ditch effluent was studied to determine its feasibility for use as a control parameter. Results observed at Bowie have shown that it is more versatile than D.O. in low rate systems.

The effluent alkalinity can be related to the alkalinity of the influent to the biological treatment system using the following relationships:

1) For an influent Total Kjeldahl Nitrogen of N_1 mg/L, of which N_2 is organic nitrogen, and N_{nb} mg/L of organic-N is non-biodegradable,

Alkalinity released during the deamination of organic nitrogen in the influent wastewater = $3.57\ (N_2 - N_{nb})$

Total alkalinity available after deamination (A_2)
= Influent alkalinity + $3.57\ (N_2 - N_{nb})$.

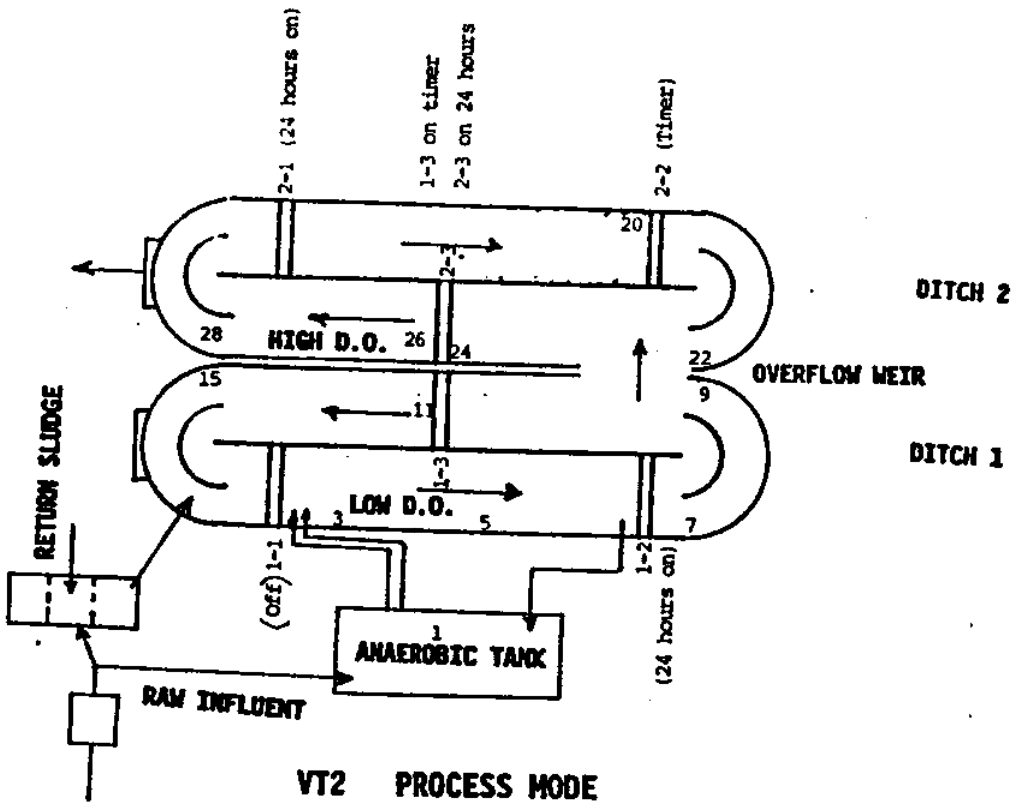


Figure 18. Dissolved Oxygen Profile in the Oxidation Ditches in the VT2 Process in August 1989

Average of measurements on 8/8, 8/18 and 8/24/89

Location No.	Description	D.O. (mg/L)
1	Anaerobic Cell (Center, 5' depth)	0.00
3	25' from rotor 1-1 (always off)	
	1 Anaerobic Pump on	0.10
	2 Anaerobic Pumps on (reaeration by anaerobic pumps when liquid drops into ditch 1)	0.15
5	50' from rotor 1-1	
	1 Anaerobic Pump on	0.00
	2 Anaerobic Pumps on	0.05
7	20' from rotor 1-2 (always on)	0.20
9	Overflow weir - in ditch 1	0.10
11	Rotor 1-3 off	0.05
	Rotor 1-3 on	0.3 - 1.0
15	Point of regular measurement in ditch 1	0.05
20	10' before rotor 2-2 (rotor 2-1 operates 24 hours/day)	0.15
22	Overflow weir, in ditch 2	0.15
24	10' before rotor 2-3	0.05
26	20' below rotor 2-3 (operates 24 hours/day)	0.6 - 1.2
28	Point of regular measurement (in ditch 2)	0.10

All D.O. measurements are made at 2 to 5 feet depth of liquid

2) If N_3 mg/L of the influent nitrogen is incorporated in the biomass as organic nitrogen (during aerobic and anoxic BOD stabilization and growth of autotrophs),

Alkalinity generated during biomass production = - 3.57 N_3.

3) Nitrogen available for nitrification = $N_1 - N_3 - N_{nb}$.

If N_4 mg/L of ammonia-N is discharged in the effluent, the alkalinity generated in nitrification = - 7.14 $(N_1 - N_3 - N_{nb} - N_4)$.

Note: Alkalinity consumption is expressed as (- generation)

4) If N_5 mg/L of nitrate-N is discharged in the effluent, nitrate available for denitrification = $(N_1 - N_3 - N_{nb} - N_4 - N_5)$.

Alkalinity generated during denitrification = 3.57 $(N_1-N_3-N_{nb}-N_4-N_5)$.

5) Effluent Alkalinity = Step 1 + Step 2 + Step 3 + Step 4

$$= \text{Influent Alk} + 3.57\,(N_2 - N_{nb}) - 3.57\,N_3 - 7.14\,(N_1 - N_3 - N_{nb} - N_4) + 3.57\,(N_1 - N_3 - N_{nb} - N_4 - N_5).$$

$$= \text{Influent Alk} + 3.57\,(N_2) - 3.57\,N_1 + 3.57\,N_4 - 3.57\,N_5 \qquad \ldots.5$$

6) If N_6 mg/L of nitrate is present in the influent because of recycle from solids handling, a term (+ 3.57 N_6) should be added to the right hand side of Step 5 to account for its denitrification.

Since [Influent Alk + 3.57 $(N_2 - N_{nb})$] is the total available alkalinity (A_2) and (3.57 N_4 - 3.57 N_5) is the effect of effluent nitrogen forms,

equation 5 can also be written as

$$\text{Eff Alk} = \text{Total Avail Alk} - 3.57\ N_1 + \text{Eff N effect} \quad \ldots 5a.$$

Therefore the effluent alkalinity is a function of:

Organic N in the influent;
Total Kjeldahl Nitrogen in the influent;
Total available alkalinity; and
Effluent ammonia, alkalinity and nitrates.

The computation is summarized in Table 10. The relationship is independent of MCRT because nitrogen removed in the sludge has the same effect on alkalinity as nitrogen removed through nitrification and denitrification. This is shown in Table 11.

The non-biodegradable organic nitrogen in the influent to the secondary treatment system will leave the plant in the waste sludge or in the effluent. Equation 5 shows that the non-biodegradable organic nitrogen does not affect the final form of the equation.

Effect of Nitrites

The alkalinity consumed across the secondary treatment system does not change if nitrite-N is discharged in place of nitrate-N. If the alkalinity consumed during the synthesis of biomass in nitrification (nitrifiers) is calculated separately (Step 2 of the model), the stoichiometry shows that all the alkalinity consumed during the transformation of ammonia to nitrite and then to nitrate is in the first step of the two step reaction. The oxidation of nitrite to nitrate does not consume any alkalinity (Sen et.al., 1990)

Denitrification of nitrite-N will generate the same amount of alkalinity as observed for the denitrification of nitrate-N when the alkalinity consumed during the production of biomass during denitrification is accounted for separately (as in Equation/Step 2 of the model). The stoichiometry of the alkalinity balance model at steady and non steady states will be published separately.

Table 10. Alkalinity Balance at a Treatment Facility

Nitrogen	mg/L	Alkalinity	mg/L as $CaCO_3$
Influent TKN	N1	Influent Alkalinity	A1
Influent Organic-N	N2	Alkalinity from Deam. Avail Alk (A2)	3.57 N2 A1 + 3.57N2
Sludge N[1]	N3	Alkalinity Generated	-3.57 N3
Effluent Ammonia-N	N4		
Nitrificaition (N7)	N1-N3-N4	Alkalinity Generated	-7.14 (N1-N3-N4)
Effluent Oxidized-N	N5		
Denitrification (N8)	N1-N3-N4-N5	Alkalinity Generated	3.57 (N1-N3-N4-N5)
		Effluent Alkalinity	A2 - 3.57 N1 + 3.57 N4 - 3.57 N5

Effluent Alkalinity = Available Alkalinity - 3.57 (Influent TKN) - Effl N effect

[1] In terms of influent flow

Note: The effect of nitrite and nitrate-N forms is discussed later.
Non-biodegradable organic N does not affect the final form of the equation.

Table 11. Effect of MCRT on Effluent Alkalinity (Constant Temperature)

Nitrogen is in mg/L as N in terms of influent flow
Alkalinity is in mg/L as $CaCO_3$ in terms of influent flow

	MCRT1 days	MCRT2 days
Influent TKN	N1	N1
Available Alkalinity	A2	A2
Sludge N [1]	N3	N3 - 1
Alkalinity Generated	-3.57 N3	- 3.57 (N3-1)
Effluent Ammonia-N	N4	N4
Nitrification [2]	N7	N7 + 1
Alkalinity Generated	-7.14 N7	-7.14 (N7 + 1)
Effluent Oxidized-N	N5	N5
Denitrification [3]	N8	N8 + 1
Alkalinity Generated	3.57 N8	3.57 (N8 + 1)
Effluent Alkalinity	A2 - 3.57 N3 - 7.14 N7 + 3.57 N8	A2 - 3.57 N3 - 7.14 N7 + 3.57 N8

MCRT2 > MCRT 1

[1]Sludge production is less at MCRT2 than at MCRT1
The calculation shows a net reduction of 1 mg/L in the amount of nitrogen removed in the biomass.

[2]The decrease in sludge production increases ammonia available for nitrification. at MCRT2.

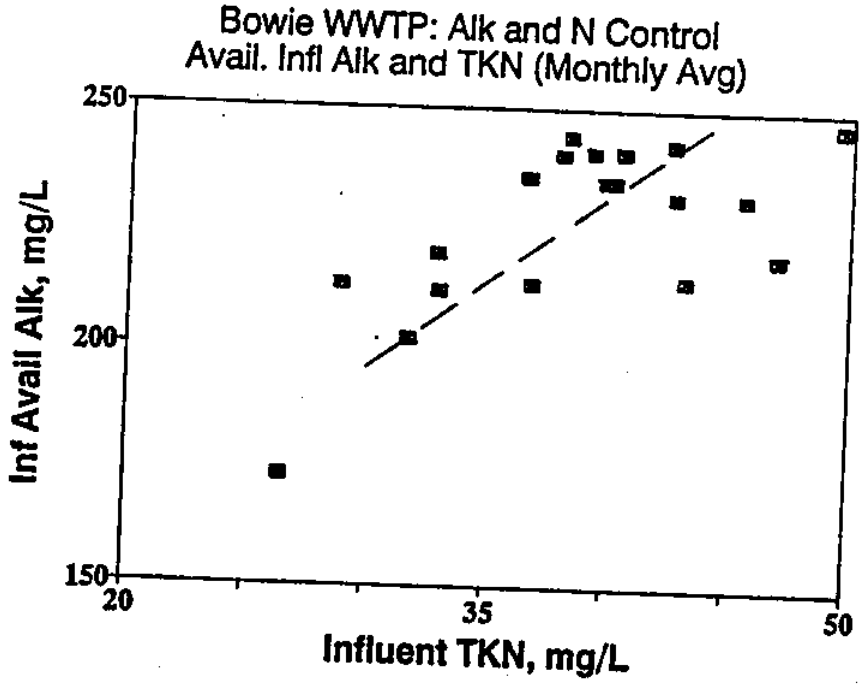
[3]Since effluent oxidized-N remains constant, denitrification increases with nitrification at MCRT2.

Precautions must be taken if equation 5 is to be applied to final effluent alkalinity at a plant where the secondary effluent is chlorinated. A short term increase in aeration effluent nitrite will result in an increase in the chlorine demand because chlorine will be consumed during the oxidation of nitrite to nitrate in the contact chamber. This will result in a drop in the final effluent alkalinity of 3.57 mg/L for each mg/L of nitrite-N oxidized by chlorine in the contact chamber.

Application Scenarios

For most municipal wastewaters, the effluent alkalinity will be independent of the concentration of nitrogen in the influent or the influent alkalinity. The influent alkalinity measured will fluctuate with the extent of deamination. However, the total available alkalinity is proportional to the influent TKN. An increase in influent TKN concentration of y mg/L results in an increase in the total available alkalinity of 3.57y mg/L. This is shown in Figure 19 where the total available alkalinity (A2) measured at the Bowie WWTP increases by 3.57 mg/L for every mg/L increase in influent TKN. The additional influent nitrogen is removed in the biomass or through denitrification, both of which have the same net effect on effluent alkalinity. The effluent alkalinity for complete nitrogen removal can be computed from equation 5a to determine an operating set point and range. The deviations from the set point will be a function of the effluent ammonia-N, oxidized-N (nitrite-N + nitrate-N) levels.

This discussion has outlined how effluent alkalinity may be used to control ammonia and oxidized-N levels in the effluent. An increase in ammonia levels of 1 mg/L as N results in an increase in effluent alkalinity of 7.14 mg/L. A reduction in the amount of nitrate-N denitrified by 1 mg/L for N reduces the effluent alkalinity by 3.57 mg/L. This is shown in Column 3 in Table 12 where the effluent ammonia-N has increased by 1 mg/L over that in the reference scenario while the nitrate-N level remains constant at N_5 mg/L (column 2). This may be typical of a temperature limitation on performance in which nitrification has been affected but denitrification has not. As a result, the effluent total nitrogen has increased by 1 mg/L.



Δ Available Influent Alkalinity / Δ Influent TKN = 3.57

Figure 19. Relationship Between Available Alkalinity and Influent TKN at the Bowie WWTP

Table 12. Partial Loss of Nitrification due to Temperature and Air Limitations on Alkalinity
(Sludge Production is maintained constant for the simulation)

All nitrogen concentrations are in terms of influent flow.

	Normal	Temperature Limitation[1]	Air Limitation[2]
Influent TKN	N1	N1	N1
Available Alkalinity	A2	A2	A2
Sludge N	N3	N3	N3
Alkalinity Generated	- 3.57 N3	- 3.57 N3	- 3.57 N3
Effluent Ammonia-N	N4	N4 + 1	N4 + 1
Nitrification	N7	N7 - 1	N7 - 1
Alkalinity Generated	- 7.14 N7	- 7.14 (N7-1)	- 7.14 (N7-1)
Effluent Oxidized-N	N5	N5	N5 - 1
Denitrification	N8	N8 - 1	N8 - 1
Alkalinity Generated	3.57 N8	3.57 (N8-1)	3.57 (N8-1)
Effluent Alkalinity	A9	A9 + 3.57	A9 + 7.14

1 This may be the predominant factor.

2 This may be the predominant factor. Temperature may also cause this change.

In column 4, the effluent ammonia-N has increased by 1 mg/L. However, the effluent nitrate decreased by 1 mg/L from that in reference column 2. This maintained the amount of nitrate-N denitrified at the same level as that in the reference scenario. The effluent total nitrogen remains constant. This may be typical of an air limitation on nitrification. In this case, the effluent total nitrogen increased by 7.14 mg/L.

With an increase in the amount of air supplied and in temperature, the changes outlined in Table 12 will be reversed. During a time period when an increase in nitrification is accompanied with an increase in denitrification, the effluent ammonia-N decreases by 1 mg/L but the nitrate-N remains constant. The total nitrogen decreases by 1 mg/L for each additional mg/L of ammonia nitrified. The increase in nitrification consumes an additional 7.14 mg/L of alkalinity for each mg/L of ammonia nitrified and the corresponding increase in denitrification recovers 3.57 mg/L. As a result, the effluent alkalinity decreases by 3.57 mg/L for each mg/L reduction in total nitrogen. This represents the upper portion of the graph shown in Figure 20, where the effluent ammonia-N (and TKN) concentrations are decreasing and the effluent nitrate-N is close to 0 mg/L. After a certain point, the capacity of the system to denitrify the nitrate-N is exceeded. Any increase in air beyond this point results in an increase in nitrate-N with further reduction in ammonia-N. This results in a consumption of 7.14 mg/L of alkalinity for each additional mg/L of nitrification and no additional recovery in denitrification. This is the lower portion of the graph in Figure 20. At the Bowie WWTP, the change in slope takes place at ammonia-N of 1 mg/L in summer and 2 mg/L in winter. The total N level where this change occurs can be interpreted as the point where the denitrification capacity of the system is exceeded. Any further decrease in ammonia-N level below that for the change in slope will result in at least an equal increase in effluent nitrate-N.

A simulation of conditions at the Bowie plant, based on actual operating conditions, is shown in Table 13. The reference effluent alkalinity for complete nitrification and denitrification was 75 mg/L. Effects of changes in effluent ammonia and nitrates, extent of deamination of organic

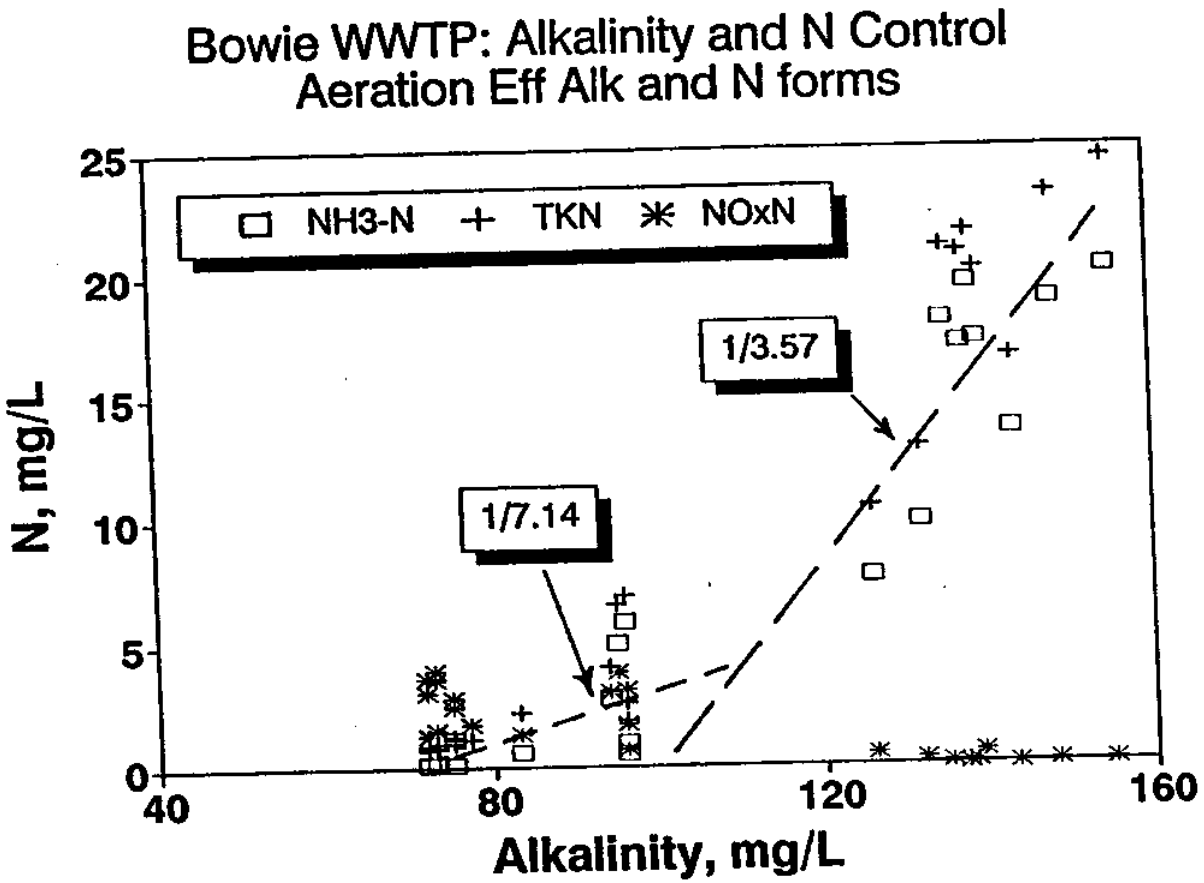


Figure 20. Effluent Nitrogen Forms and Alkalinity at the Bowie WWTP

Table 13. Simulation of Effluent Alkalinity Conditions at Bowie WWTP.

	Combined Influent				Secondary Effluent				Sldg N	DN	Eff TN
	Alk	TKN	Org-N	COD	NH_3	NO_3	Alk	^ Aern			
	----- mg/L ----				-- mg/L --			inches	--- mg/L ---		
Optimum[1]	146	30	10	500	0	0	75	0	10	20	0
Air Limit	146	30	10	500	1	0	78.5	+0.25	10	19	1
Air Excess	146	30	10	500	0	1	71.5	-0.25	10	19	1
Less Deam.	140	30	12	500	0	1	71.5	-0.25	10	19	1
High Deam.	153	30	8	500	0	1	71.5	-0.25	10	19	1
High Conc	161	32	8	560	0	1	71.5	-0.25	12	19	1
High MCRT	161	32	8	560	0	1	71.5	-0.25	10	21	1
Aerobic MCRT not adequate for temperature - experienced in December 1988											
Case 1	146	30	10	500	5	0	93	0	10	15	5
Aerobic and anoxic MCRTs are both inadequate for temperature - December 1988											
Case 2a	146	30	10	500	2	2	75	-0.25	10	16	4
Case 2b	146	30	10	500	5	5	75	-1.25	10	10	10
Case 2c	160	30	10	500	7	3	89	-0.75	10	10	10

[1] Effluent Alkalinity = Available Influent Alk (A_2) - 3.57 (Influent TKN) + Effluent N effect

When effluent TN = 0
Effluent Alkalinity = Available Influent Alkalinity - 3.57 (Influent TKN).

Under Optimum Conditions
Effluent Alkalinity = 146 + 3.57 (10) - 3.57 (30)
= 75

nitrogen, influent alkalinity and MCRT on the secondary effluent alkalinity have been computed. The influent characteristics and nitrogen removal in the waste sludge are based on actual parameters observed in the VT2-BNR process. This BNR process was operated at MCRTs which were adequate for nitrification at the liquid temperatures experienced. The results of the simulation may be compared to the results for the VT2-BNR process presented earlier in Tables 4 (ditch alkalinity) and 7 (effluent nitrogen forms). The recommended changes to the aeration are expressed in terms of changes in the depth of liquid in the oxidation ditch. This controls the depth to which the brush aerators are submerged.

The liquid level was adjusted when the effluent alkalinity measurements were outside an operating range of 72 and 75 mg/L. For every 3 mg/L deviation from this range, the weir was adjusted by 0.25 inch. The timers on aerators were adjusted if the alkalinity deviated 6 or more units outside the normal operating range. Appropriate adjustments were made prior to weekends during which high flows were expected. Consideration was also given to weather forecast to fine tune some changes. Changes had to be made more frequently for the brush aeration system because the amount of air transferred is more sensitive to changes in atmospheric temperature than a system which uses diffused aeration (where the air supplied by blowers). Liquid temperatures are more sensitive to changes in atmospheric temperature because of turbulence and rapid heat loss at the liquid surface. The atmospheric temperature also affects the partial pressure of oxygen in the saturated layer of air above the oxidation ditch.

Two cases are listed at the end of Table 13. Case 1 shows a loss of nitrification due to inadequate MCRT. This is experienced during cold weather at several high to moderate rate (4 to 12 hour nominal hydraulic retention time) activated sludge plants where secondary clarifier capacities may be a limiting factor. Effluent alkalinity levels increased because of incomplete nitrification. A similar situation was experienced at the Bowie WWTP in December, 1988, when the plant operated with one oxidation ditch only.

Three scenarios are included for the Case 2. They may be observed when attempts are made to increase the aerobic MCRT in winter by increasing the aerobic volume. The simulation presented in scenario 2c is occurs frequently because denitrification is less sensitive to temperature than nitrification. However, partial loss of anoxic volume at higher operating D.O. levels may show results similar to those presented for scenario 2a. Recycle of dissolved oxygen into the anoxic zone (cells) decreases the pounds of biomass under anoxic conditions. Scenario 2b may be interpreted as an extreme example of scenario 2a. It may be characteristic of oxidation ditches which operate with high internal recycle rates and any increase in dissolved oxygen in winter can have a significant effect on the anoxic volume.

A simultaneous increase in effluent ammonia-N and oxidized-N as simulated in scenario 2b is not indicated by the effluent alkalinity. The analysis for scenario 2b in an alkalinity balance model is presented in Table 14. Fortunately, there are a number of secondary parameters which can indicate simultaneous loss of nitrification and denitrification. These include:

a) Dissolved oxygen levels in the oxidation ditch / aeration tank;
b) Secondary effluent or ditch effluent supernatant ortho-P
c) Secondary effluent or ditch effluent supernatant turbidity, and
d) Volatile solids fraction.

A partial loss of nitrification due to a drop in temperature results in a drop in the oxygen demand. If the oxygen supply is not reduced, it may result in an increase in the concentration dissolved oxygen. Recycle of oxygen will reduce the anaerobic volume available for substrate storage in the mechanism for biological P removal. Effluent ortho-P levels will increase as a result. Turbidity may increase because the mode of substrate storage and stabilization is affected.

The aeration will have to be reduced over a short period of time to recover the anaerobic volume fraction. This is the basis for the "change in aeration" in Table 13. As phosphorus removal recovers, MCRT should be increased to increase the aerobic and anoxic MCRTs. An improvement in

Table 14. Effect of Simultaneous Increase in Effluent Ammonia and Oxidized-N forms on Alkalinity
(Sludge Production is maintained constant for the simulation)

All nitrogen concentrations are in terms of influent flow.

	Normal Condition	Process Upset Condition
Influent TKN	N1	N1
Available Alkalinity	A2	A2
Sludge N	N3	N3
Alkalinity Generated	- 3.57 N3	- 3.57 N3
Effluent Ammonia-N	N4	N4 + 1
Nitrification	N7	N7 - 1
Alkalinity Generated	- 7.14 N7	- 7.14 (N7-1)
Effluent Oxidized-N	N5	N5 + 1
Denitrification	N8	N8 - 2
Alkalinity Generated	3.57 N8	3.57 (N8-2)
Effluent Alkalinity	A9	A9

denitrification will increase the effluent alkalinity. The air supply should be increased when the alkalinity increases above the normal operating range. Subsequent increase in alkalinity may result in an oxygen limitation on nitrification. If phosphorus removal and turbidity deteriorate when aeration is increased, the total MCRT is not adequate and should be increased further.

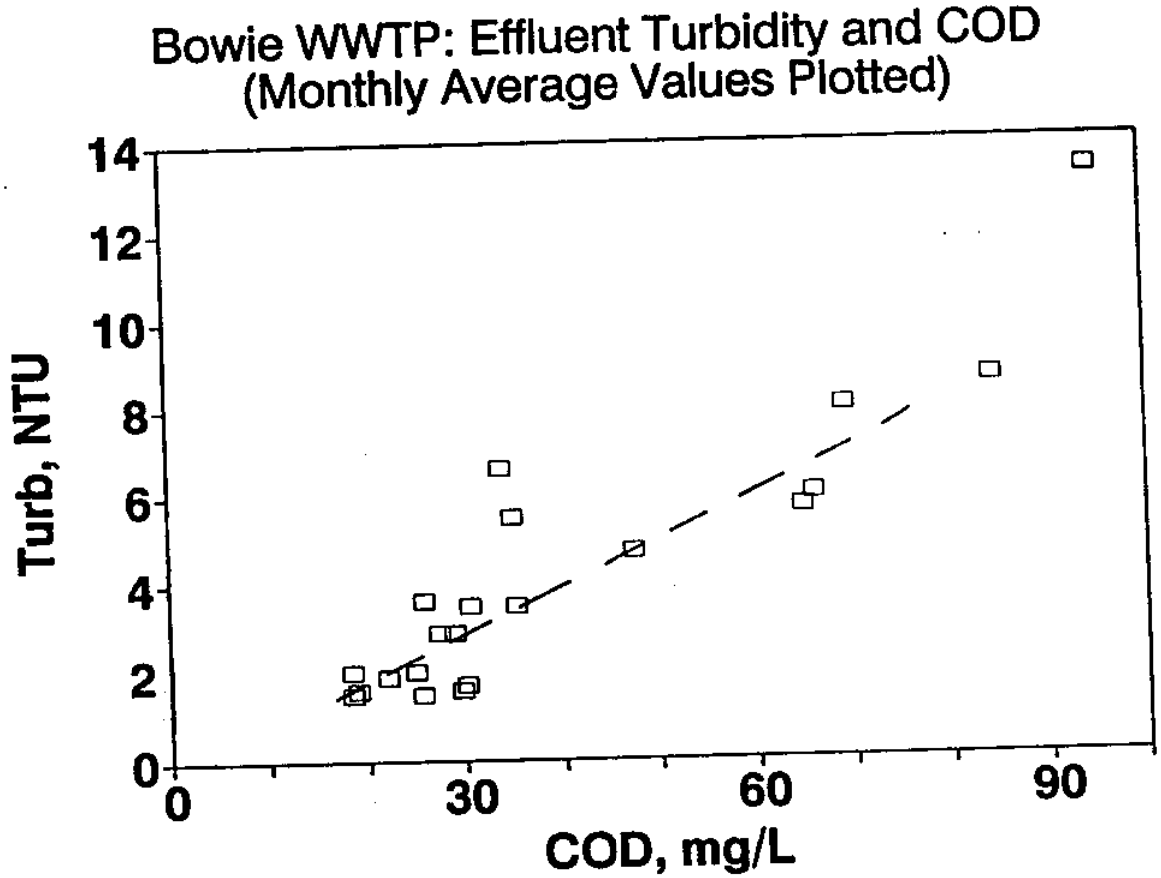
Turbidity as a control parameter

Turbidity can be used as an effective control parameter for phosphorus and nitrogen removal in low rate systems operating at high nitrate recycle rates. When aeration is less than, turbidity tends to increase because of one or more of the following:

a) soluble COD is not reduced to levels which may be observed during efficient treatment; and
b) particulate matter is not captured efficiently in the bacterial floc because of reduced extracellular enzymatic activity.

Monthly average values for effluent COD and turbidity from the Bowie WWTP are plotted in Figure 21. Total COD measurements are used in this figure because turbidity is a composite of the clarity of the liquid phase (affected by soluble COD) and the particulate phase (affected by particulate COD). During Phase II, the turbidity levels increased when oxygen availability limited COD removal. Turbidity levels improved from Phases III to V as better control over aeration was attained and effluent CODs dropped below 20 mg/L (Table 7).

During biological phosphorus removal, the increase in effluent turbidity was reflected in an increase in effluent ortho and total phosphorus. In Figure 22, the effluent phosphorus is plotted against effluent turbidity for Phases III to V (after chemical phosphorus removal was discontinued). The results show that turbidity levels reflect extent of phosphorus removal in a system in which aeration for BOD removal and nitrification are optimized to maximize denitrification and biological P removal.



Note: The dashed line does not show a mathematical relationship. It is indicative of a trend.

Figure 21. Final Effluent Turbidity and Secondary Effluent Turbidity at the Bowie WWTP

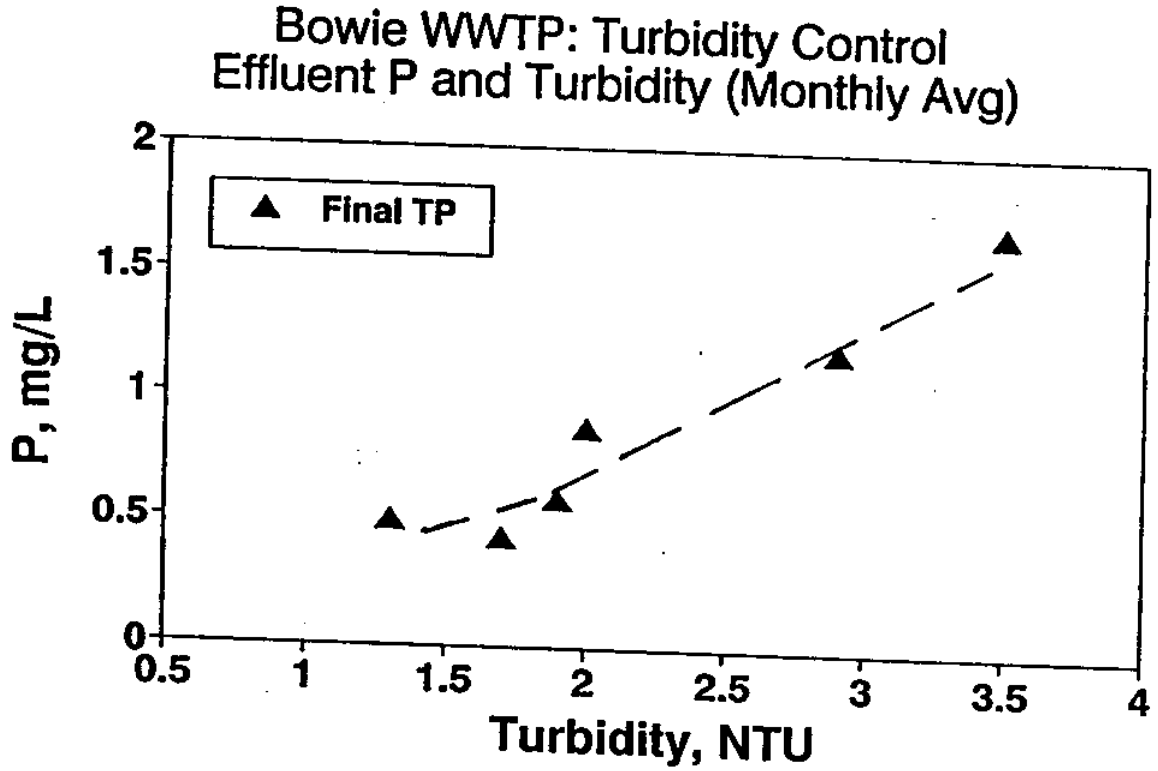


Figure 22. Final Effluent Total Phosphorus and Secondary Effluent Turbidity

An increase in turbidity due to insufficient aeration is observed in most activated sludge systems. However, to utilize it as a control parameter, changes have to be made in response to small increases in turbidity to prevent an increase in ammonia levels in the effluent. Turbidity control may be easier in certain low rate systems with low F/M ratios and high recycle rates, than in high rate systems containing CSTRs in series such as the A^2/O and VIP processes.

Low Nitrate Recycle, High Rate System

In an activated sludge system with low recycle rates and multiple aerobic cells, turbidity decreases rapidly in the first cell where heterotrophic bacteria stabilize soluble and stored forms of BOD. Nitrification is inhibited in the first cell unless sufficiently high levels of oxygen can be maintained or biofilms dominated by nitrifiers are placed in contact with dissolved oxygen. A nitrifier located in the vicinity of a heterotrophic bacteria inside a bacterial floc is starved of oxygen and ammonia. As a result, the effluent from the first cell contains a high concentration of ammonia-N. However, if most of the BOD is stabilized in this cell, the turbidity (of the settled mixed liquor) from this cell will be close to that in the clarified effluent.

In the following aerobic cell, heterotrophic growth rate is limited by the lack of available BOD. Nitrifiers grow readily in the presence of available oxygen and are the principal contributors to the observed oxygen uptake rate. However, the change in turbidity across this cell may not be significant.

A significant increase in effluent turbidity may result in a substantial loss of nitrification because this would indicate the failure of the first aerobic section to supply enough air, extending its reactions into the second aerobic section. Therefore, it may be difficult to control nitrogen and phosphorus removal unless turbidity is measured in the first aerobic cell to control aeration to this cell.

High Nitrate (or Internal) Recycle, Low Rate System

In an activated sludge system operating with a high nitrate recycle, the influent COD is diluted extensively by the recycle. This can reduce the heterotrophic growth rate and increase the growth rate of nitrifiers in the first cell. If the system is operated at a low F/M, a higher fraction of biomass is composed of nitrifiers. This may also increase the extent of nitrification in the first cell. Therefore, BOD stabilization, P uptake, nitrification and the reduction in turbidity tend to occur simultaneously. A significant increase in effluent turbidity affects nitrogen and phosphorus removal but does not cause a radical as that in the low recycle system. Therefore, turbidity control of nitrogen and phosphorus removal is more effective as the recycle rate is increased.

At the Bowie WWTP, it was observed that high flows on weekends were followed by higher effluent turbidities, alkalinities and phosphorus levels in the secondary effluent on Monday unless the aeration was increased to compensate for the increase in loadings. The actual HRT of the oxidation ditch and clarifier system is 20 hours. An example of a short term increase in alkalinity following a weekend is shown in Figures 23 and 24. On the average, an increase in turbidity of 0.5 NTU resulted in an increase in effluent ortho-P of 1 mg/L, effluent alkalinity of 10 mg/L and effluent ammonia of 2 mg/L in Phase V (VT2-BNR process).

The aeration can be controlled by turbidity by setting up a low and a high level for an operating turbidity range. The air supplied is increased if the turbidity exceeds the high level. It is reduced when the turbidity drops to the low level. This principle is used in the D.O. Minimizer™ developed by Schreiber Corporation and used with Schreiber Counter Current diffusers (Figure 25). These tanks operate at high MCRTs and with high internal recycle rates. The internal recycle (IR) in the counter current diffuser system can be calculated as follows:

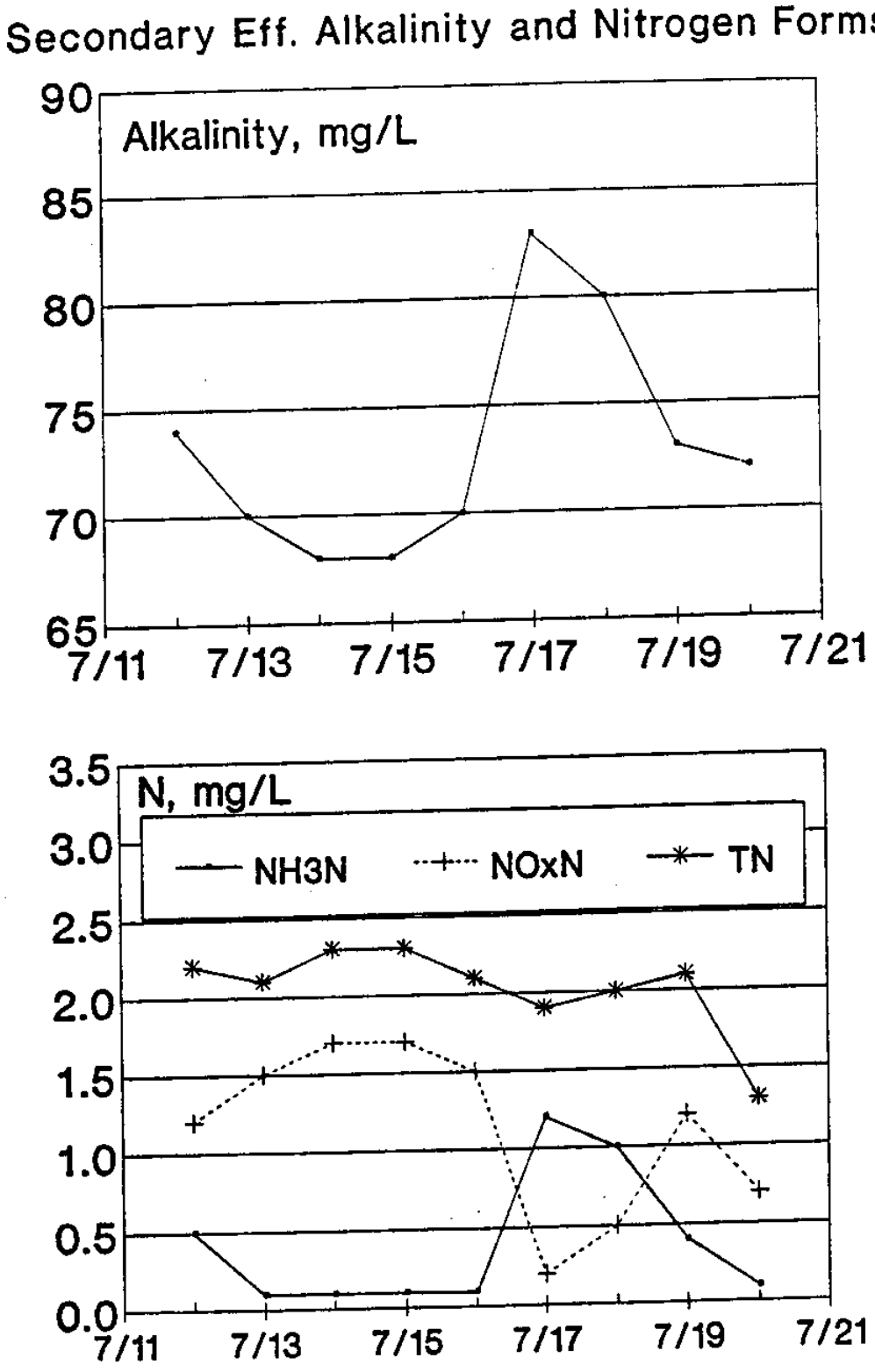


Figure 23. Secondary Effluent Alkalinity and Nitrogen Forms, July 1989

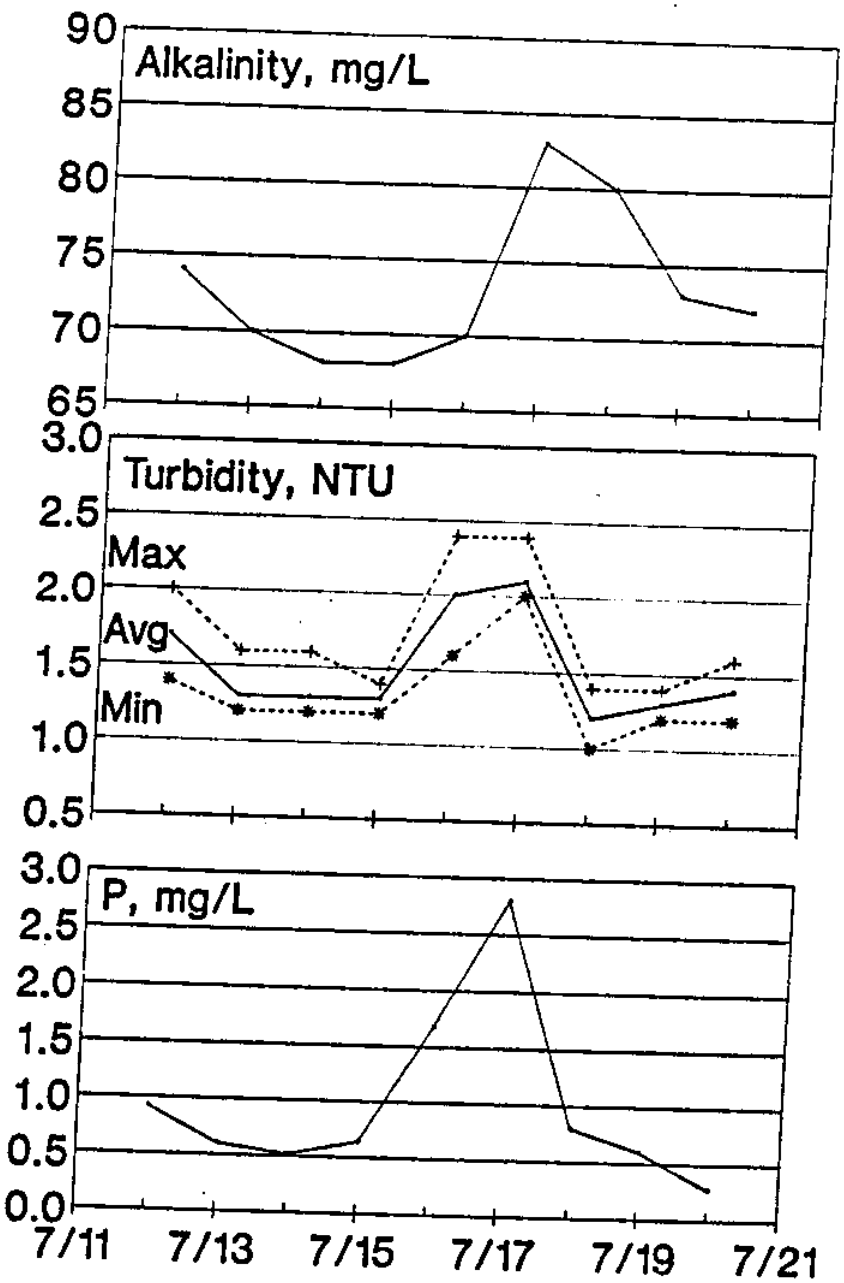


Figure 24. Secondary Effluent Alkalinity, Turbidity and Phosphorus Forms, July 1989

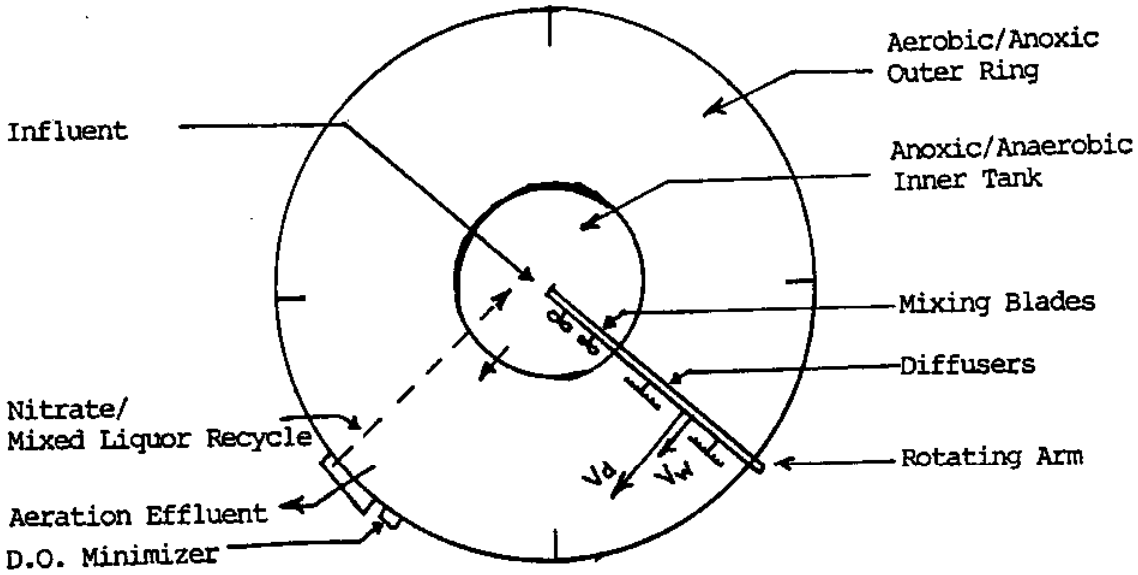


Figure 25. Schematic of Schreiber Counter Current Diffuser System™

$$IR = (v_d - v_w) * A / Q$$

where

v_d is the tangential velocity of the diffuser arm at the mid-point (radially) of the outer ring,

v_w is the tangential velocity for water at the mid-point of the outer ring

A = Cross sectional area of the outer ring

Q = Raw Influent Flow Rate

The internal recycle rates are in excess of 200Q in the Schreiber Counter Current Diffuser Process at 1 day nominal HRT (Maryland City WRF, 1990). The high internal recycle makes the Schreiber Counter Current Diffuser Process an excellent candidate for turbidity control for nitrogen and phosphorus removal.

Volatile Solids and Nitrogen Removal

The volatile solids fraction increases with the loss of nitrification. The volatile solids fraction decreases with an increase in the rate of endogenous decay. Endogenous decay rates are higher under aerobic conditions than under anoxic and anaerobic conditions. Therefore, a reduction in the aerobic volume due to air limitation reduces the volume weighted average rate of endogenous decay.

An increase in volatile solids fraction during a time period when the liquid temperatures are increasing is reflected in a loss of nitrification. Temperature increases the rate of endogenous decay. Therefore, if the volatile solids fraction should decrease with temperature, it is indicative of air limitation. This trend has been confirmed at the Patuxent WRF (1990).

Phosphorus Release in the Clarifiers

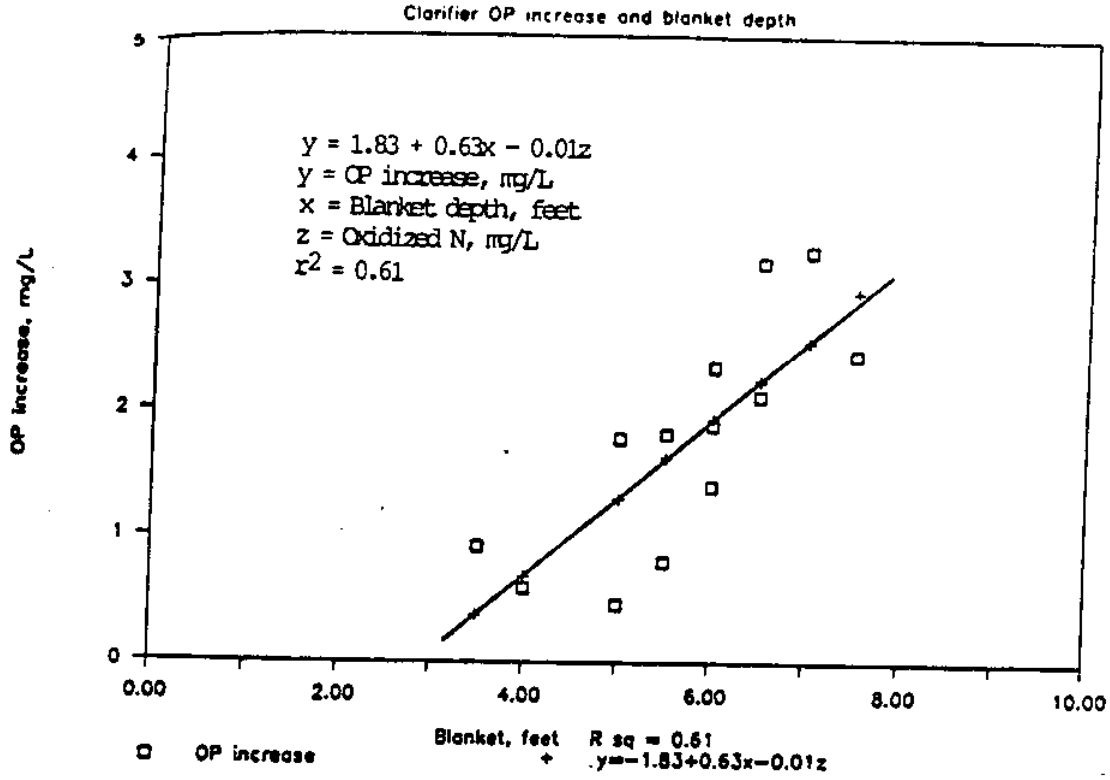
Analysis of results from Phase III (during which the plant was operated with only one oxidation ditch) showed significant levels of phosphorus release across the secondary clarifiers. In the rim feed clarifier system, where the flow is introduced at the rim, one foot off the bottom of the side wall of the clarifiers, the secondary effluent has to travel through a part of the sludge blanket (Figure 3). Therefore, some of the phosphorus which was released into solution in the sludge blankets was carried out in the secondary effluent.

The phosphorus release correlated well with the depth of the sludge blanket in Phase III (Figure 26). However, it did not appear to have any correlation with oxidized-N concentrations below 2 mg/L.

After the facility switched over to operation with two oxidation ditches in series, phosphorus release in the clarifiers dropped to very low levels, even at moderate sludge blanket depths. The soluble COD level in the effluent dropped from an average in excess of 30 mg/L in Phase III to 20 mg/L during Phase V (Table 7). Though it was difficult to determine the concentration of non-biodegradable COD since soluble BOD concentrations were not measured, the lowest concentration of soluble COD recorded was 9.95 mg/L in July 89. The BOD_5 levels dropped from 18 mg/L in Phase III to 3 mg/L in Phase V.

Phosphorus release may occur under anaerobic conditions in the secondary clarifier sludge blankets when readily available forms of COD are present in the aeration effluent. Between 10 to 15 mg/L of biodegradable soluble COD was present in the ditch effluent in Phase III but dropped to 0 to 5 mg/L in Phase V. The increase in MCRT for the same level of MLSS in Phase V (because of a doubling in the volume of mixed liquor) reduced oxygen and nitrate uptake rates (by about 40 percent for the same MLSS).

These changes reduced the rate of release of phosphorus in the sludge blankets. It was not necessary to reaerate the aeration effluent before



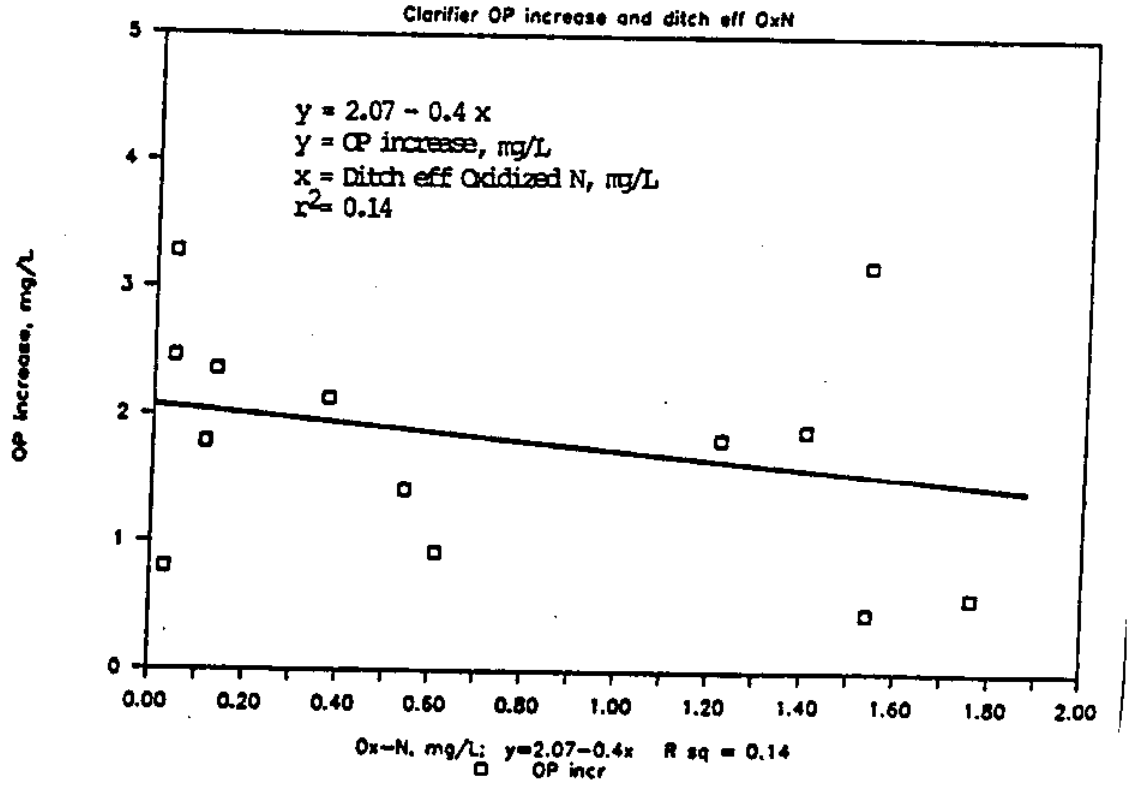


Figure 26. Graphs of Phosphorus Release Across the Secodary Clarifiers plotted against Blanket Depth and Ditch Effluent Oxidized-N

the secondary clarifiers. No significant release was measured across the clarifiers at sludge blanket depths of five feet.

Phosphorus and Nitrogen Recycle From Solids Handling Units

Approximately 6 percent of the phosphorus removed in the secondary treatment system was recycled to the headworks during gravity thickening of waste activated sludge. Precautions were taken to minimize solids recycle with the supernatant from the gravity thickeners.

Average volume of supernatant = 30,000 gpd
Concentration of phosphorus in supernatant = 30 mg/L
Influent flow = 2.2 MGD
Contribution to combined influent phosphorus concentration
$= (30{,}000 * 30)/(2.23 * 10^6)$
= 0.4 mg/L.

An additional 14 percent of the phosphorus removed during secondary treatment was recycled with the filtrate and the washings from the belt presses (contributing 0.95 mg/L to the combined influent phosphorus). These recycles increased the influent phosphorus concentration from 5.6 mg/L to 7.0 mg/L. Biological phosphorus removal removed 6.5 mg/L of the phosphorus.

Between 10 and 15 percent of the nitrogen removed in the waste sludge was recycled from the solids dewatering units. Since only 10 mg/L of the influent TKN, is removed in the waste sludge (20 mg/L was removed by denitrification), the recycle increased the influent nitrogen concentrations by 1.5 mg/L only. Therefore, nitrogen recycle is not as much of a concern as phosphorus recycle from solids handling units.

A the key finding during the full scale retrofit modification has been that gravity thickeners can be sequenced with activated sludge units used for enhanced biological phosphorus uptake if the influent COD/P ratio is still adequate after the additional phosphorus load imparted by the recycle. In

parts of the U.S. where a phosphate detergent ban has been implemented, this ratio should be adequate.

Analysis of Filamentous bacteria and SVIs

The population of filamentous bacteria was monitored on a day to day basis at the Bowie WWTP by observing samples under a microscope and measuring changes in Sludge Volume Index (SVI). The population of filamentous bacteria and SVI increased with the liquid temperature if the aeration was not regulated to maintain strict control over aerobic and anoxic volumes.

The data from 1985 shows that monthly average SVIs correlated well with liquid temperature (r^2 = 0.8) [Figure 27]. However, the correlation was not as strong in 1986 (Figure 28). To understand the conditions which result in an increase in filamentous bacteria, the daily changes in SVI, alkalinity (and therefore, nitrogen forms and soluble BOD) and temperature were studied closely. During Phase V, it was observed that the population of filamentous bacteria could be controlled if excellent nitrification, denitrification and COD stabilization were maintained. A sudden loss of nitrification and an increase in effluent soluble COD due to deficit in the amount of air supplied at warmer liquid temperatures resulted in an increase in the population of filamentous bacteria (Figure 29).

Filamentous bacteria gain a competitive advantage if there is a lack of dissolved oxygen and a relative abundance of products from cell decay (Jenkins et.al. 1986). Growth and decay rates of bacteria increase with temperature. This increases the concentration of the products from cell decay. Between May and October, a short term deficit (12 to 36 hours) in aeration (reflected in an increase in effluent alkalinity in Figure 29) resulted in an increase in SVIs. The SVIs decreased over a period of time when proper aeration and control volumes were maintained.

The results have shown that it is necessary to maintain a strict control over the relative volumes under aerobic, anoxic and anaerobic conditions. Lack of flow pacing, sudden loss of part of the aeration, or less than

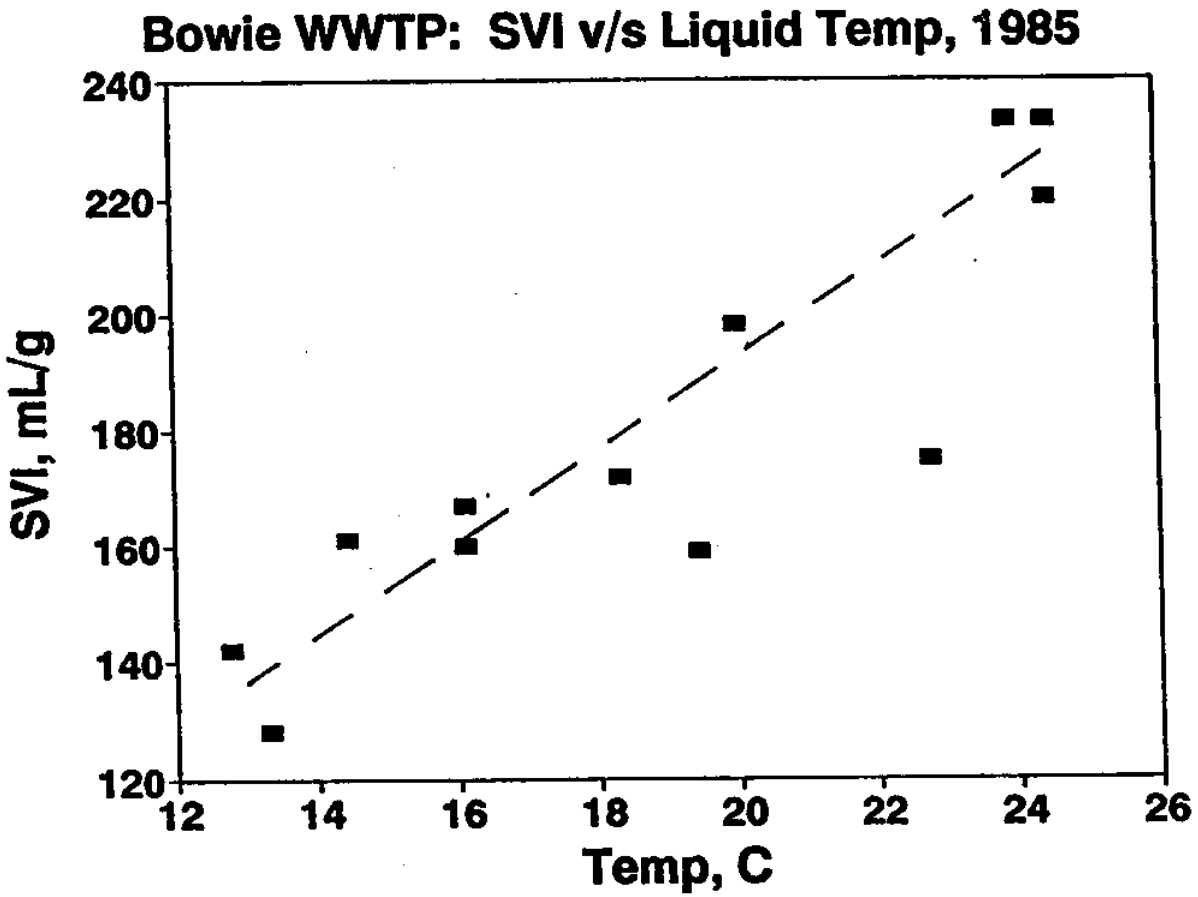


Figure 27. Sludge Volume Index (SVI) and Mixed Liquor Temperature, 1985

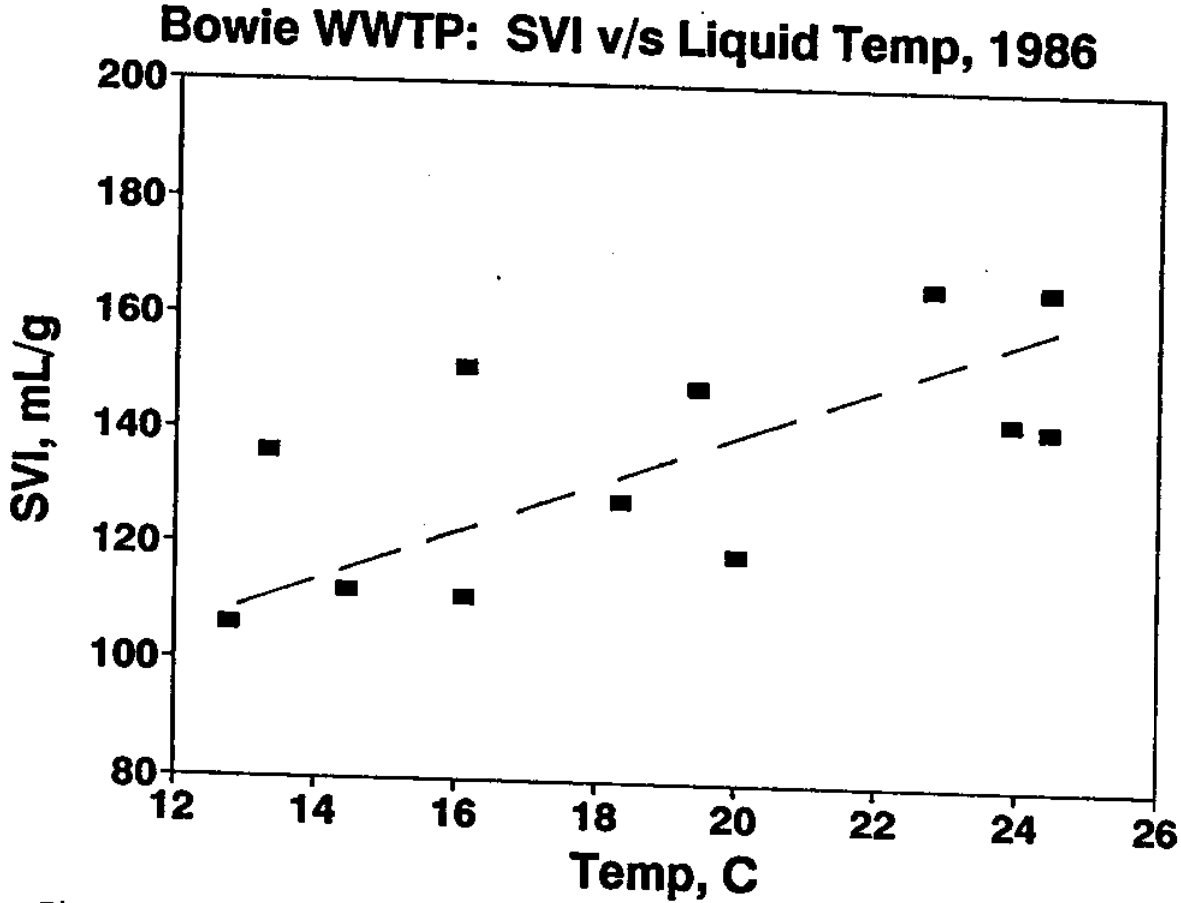


Figure 28. Sludge Volume Index (SVI) and Mixed Liquor Temperature, 1986

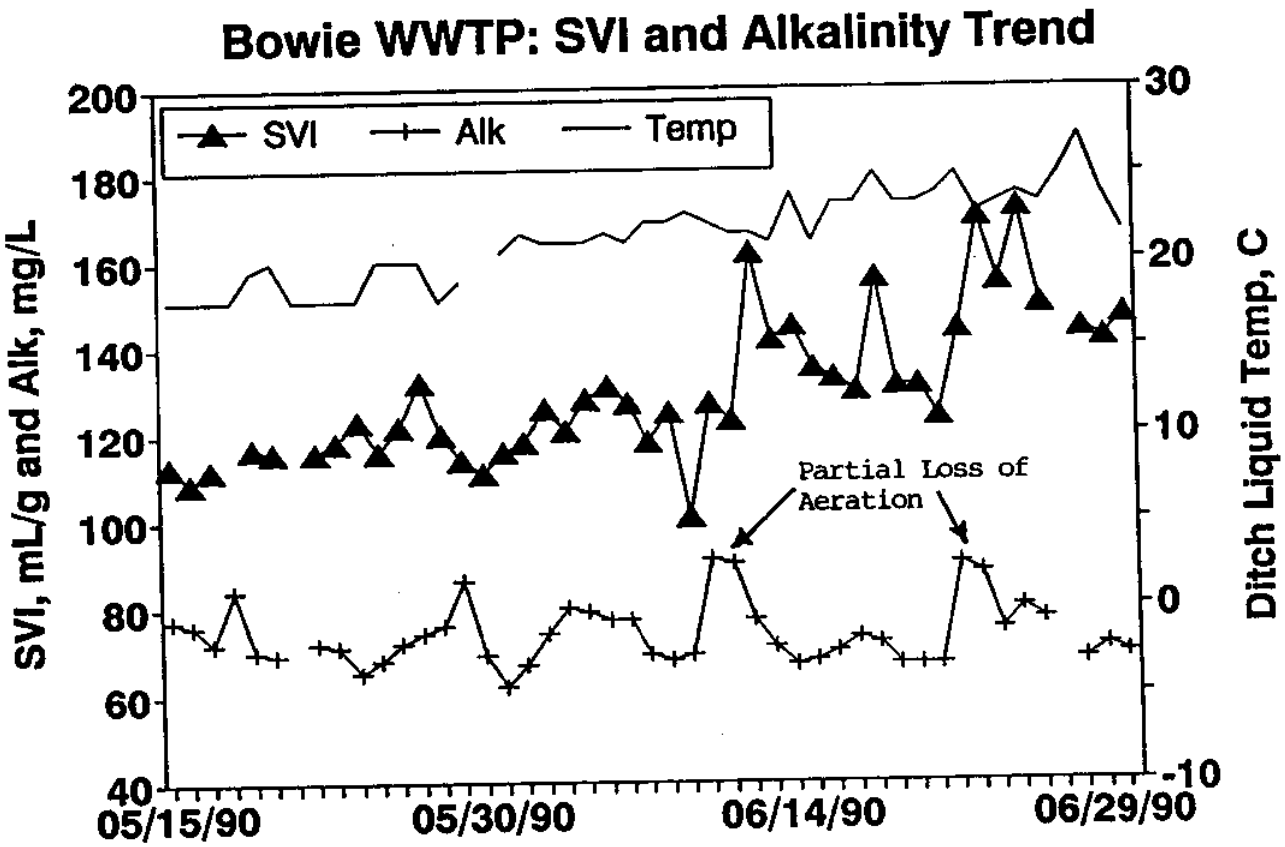


Figure 29. SVI, Alkalinity Trend and Process Disturbances, 1990

adequate temperature compensation can affect the fractions of biomass under aerobic, anoxic and anaerobic conditions. If the disturbance is large enough (in magnitude and in time) to affect the mode of COD stabilization, it will result in an increase in the population of filamentous bacteria. The observations have been confirmed during comparative studies on barrier oxidation ditches at the Patuxent and Broadneck Water Reclamation Facilities in Maryland. The Patuxent WRF has excellent automated D.O. control and low SVIs (110 mL/g) at 12 day MCRT. The Broadneck WRF lacks automated control, suffers from large fluctuations in the aerobic and anoxic volumes over a 24 hour period, and operates at SVIs above 200 mL/g at 15 day MCRT.

Oxygen and Nitrate Uptake Rates and Anaerobic Volume in Oxidation Ditches

The specific oxygen and nitrate uptake rates were very low in the oxidation ditch system at Bowie because of the long nominal HRT in the system (36 hours) and high internal recycle rates in each ditch. The specific oxygen uptake rates were the lowest in ditch 2 in which the soluble BOD_5 levels measured less than 4 mg/L. The specific susbstrate utilization rate (q) for COD stabilization is affected by the concentration of available COD as follows:

$$q = q_m \, [COD/ \, K_s + COD] \, [D.O./(K_{D.O.} + D.O.)]$$

The reported values for aerobic COD stabilization are as follows -

q_m = maximum specific substrate utilization rate, 11.2 COD/mg VSS-day
and K_s = half saturation constant, 60 to 80 mg/L as COD.

Under anoxic conditions, the values are:

q_m = 6.7 mg COD/mg VSS-day,
and K_s = 60 to 80 mg/L as COD.

These values of q_m and K_s were measured at 24 to 26 Celsius (Hanaki et. al., 1990, McClintock et.al., 1988). The maximum specific substrate

stabilization rates may decrease by 40 percent at 10 Celsius (Nutrient Control Manual, WPCF, 1983).

The specific substrate utilization rate in the second oxidation ditch will be between 5 to 10 percent of the maximum rate. The specific oxygen uptake rate is a reflection of the specific substrate utilization rate because oxygen is consumed during substrate stabilization under aerobic conditions. A number of observations are listed in Table 15.

Analysis conducted in March, 1990, showed that the nitrate uptake rate at 16 Celsius and 3000 mg/L of volatile suspended solids (MCRT of 30 days approximately) was 0.035 mg/L/min in ditch 2. In ditch 1, the nitrate uptake rate for the same temperature and volatile suspended solids measured 0.0375 mg/L/min. In the anaerobic cell, the nitrate uptake rate was 0.06 mg/L/min at a volatile suspended solids concentration of 1600 mg/L at a temperature of 16 Celsius.

At an average nitrate uptake rate of 0.035 mg/L/min and a volatile solids concentration of 3000 mg/L in the oxidation ditches (total volume of 3.3 MG), denitrification of 20 mg/L of oxidized nitrogen (generated during treatment of 32 mg/L of influent TKN) at an influent flow of 2.2 MGD requires

$$\{ (2.2 * 20) / (0.035 * 3.3 * 1440) \} * 100$$

= 26 percent of the volume of the oxidation ditches.

Anoxic stabilization of COD during denitrification of 20 mg/L of the influent nitrate will consume (20 * 2.8 mg/L = 56 mg/L) of COD. If the influent COD averages 450 mg/L, 394 mg/L has to be stabilized under aerobic conditions (assuming no anaerobic stabilization of BOD). This will generate approximately 90 mg/L of biomass at an aerobic MCRT of 15 days.

$$\text{Biomass produced} = 394 * Y_{Hmax} / (1 + k_d * 15)$$

Table 15. Specific Oxygen and Nitrate Uptake Rates at the Bowie WWTP

Date	Temperature	Anaerobic Cell		Ditch 1		Ditch 2		
		SOUR	NOUR	SOUR	SNUR	SOUR	SNUR	Ditch 2 MLVSS
	Celsius	----------		day^{-1}	----------			mg/L
8/08/89	23					0.204		1900
8/18/89	23.5			0.232		0.153		1930
8/25/89	23			0.204		0.203		1770
1/22/90	15			0.12				3200
3/11/90	16	0.256	0.052	0.131	0.016	0.102	0.015	3490

where

Y_{Hmax} = maximum heterotrophic yield = 0.50 mg VSS/ mg COD

k_d = 0.08 /day at 16 Celsius

Biomass produced = 90 mg/L VSS

COD in biomass = 90 * 1.42 = 128 mg/L

A total of 128 mg/L of COD will be removed in the biomass.

This leaves (394 - 128 = 267) mg/L to be stabilized under aerobic conditions. At an oxygen uptake rate of 0.3 mg/L/min at an MLVSS of 3000 mg/L, aerobic stabilization of 267 mg/L of COD and oxidation of 22 mg/L of ammonia will require

[[2.2 * (267 + 4.57 * 22)] / (0.3 * 3.3 * 1440)] * 100

= 56 percent of the volume in the oxidation ditches.

The remaining 18 percent of the total oxidation ditch volume (100 - 26 -56) can be maintained under anaerobic conditions. A majority of the anaerobic volume is in oxidation ditch 1. It helps enhance biological phosphorus removal in the oxidation ditches.

To increase the nitrate uptake rate (denitrification rate) and reduce the volume required for denitrification, it will be necessary to increase the concentration of available COD. Denitrification may be conducted at a higher rate in a separate anoxic tank outside the oxidation ditch if the raw influent is not diluted by a large nitrate recycle. However, the effluent oxidized-N concentration cannot be reduced to the concentrations observed during efficient treatment in oxidation ditches unless the nitrate recycle rate is increased above 300 percent. (A high recycle rate will reduce the concentration of available COD, decrease the rate of denitrification and increase the volume required for denitrification.)

Retrofit Modification and Operating Costs

Cost for the Modification

The total expenditure for the modification of the Bowie WWTP will be $400,000, approximately. This excludes process development, process engineering and start-up costs which were provided by the University through a separate grant program.

Though the facility is rated at 3.0 MGD, the available air supply, clarifier and solids handling units will not be able to handle the loadings at flows in excess of 2.5 MGD.

Calculation of Net Savings in Annual Operating Costs

The cost savings from biological phosphorus and nitrogen removal are calculated for two categories. The first category can be observed directly by comparing budgets for Phase I and Phase V. The second category was calculated for the expenditure for supplemental alkalinity and energy if the performance results for Phase V were attained with chemical phosphorus removal and denitrification reduced oxidized-N to 6 mg/L.

Direct Savings

Discontinuation of ferrous sulfate for chemical P removal results in savings of $30,000 per year;

Other (simulated or indirect) Savings

1) Discontinuation of ferrous sulfate generated an annual savings of $37,500 in supplemental alkalinity which would be added as caustic soda. Iron was added at the rate of 10.9 mg/L as Fe^{2+} to the combined influent. Precipitation of iron as ferric hydroxide or ferric phosphate consumes 30 mg/L of alkalinity. To recover this alkalinity, the addition of 30 percent caustic soda at 70 cents per gallon (1989 costs) will cost $37,500.

2) During Phase I, an average of 58.85 dry tons of sludge (398.7 wet tons) were generated each month (between October 1986 and September 1987). This includes 7 dry tons of iron sludge each month. The dry ton value is calculated after aerobic digestion in ditch 2. The total suspended solids concentrations in the final effluent averaged 15 mg/L and is included in the dry ton per month number (Table 16).

During Phase V, an average of 56.98 dry tons of sludge were generated each month between June 1989 and May 1990. This excludes dry solids added as lime. The flow and loadings have not changed since 1986. The final effluent suspended solids averaged 4 mg/L. The contribution of effluent suspended solids has been included (Table 16).

Several factors need to be considered to compare sludge production in Phase I with that in Phase V. Phosphorus was removed chemically in Phase I but was removed biologically in Phase V. Aerobic digestion of some of the waste sludge helped decrease the quantity of sludge in Phase I. The amount of iron added to remove phosphorus to levels below 1 mg/l was not large because a phosphate detergent ban reduced phosphorus levels in the influent by 40 percent. During Phase V, the plant was operated without aerobic digestion but at a higher MCRT. The net effect was that sludge production numbers did not change significantly. However, the data has to be examined in conjunction with the energy requirements.

During Phase I, 2.5 brush aerators were operated in the oxidation ditch utilized which was used for secondary treatment. At least one brush aerator was operated in the second oxidation ditch to digest waste sludge aerobically. The energy calculations for Phase I and Phase V are shown in Table 17. The aeration energy was not metered during Phase I. Therefore, the aeration energy consumption from September to December 1988 in Phase III, when only one oxidation ditch was operated for secondary treatment, was used to simulate the consumption for the corresponding period in Phase I. To determine a ratio of year round aeration energy cost to the seasonal cost from September to December, the ratio of energy costs from September to December, 1989 (Phase V) to the annual cost for Phase V was used as the conversion factor.

Table 16. Comparison of Sludge Production in Phase I and Phase V

	Phase I		Phase V	
	Per month dry ton	Per month wet ton	Per month dry ton	Per month wet ton
Dewatered Sludge	55.04	398.7	55.97	
Solids in effluent	3.84		1.01	
Dry Solids Production	58.88		56.98	
Operating Characteristics				
MCRT (days)	8 to 12		15 to 30	
Aerobic Digestion	Yes		No	
Fe ppt. sludge	Yes (included above, avg of 7 dry tons)		No	
Lime Stabilization	No		Yes (avg of 15 dry and 15 wet tons, not included above)	
Effluent NH_4-N (mg/L)	>10		0.2	
Effluent Nitrates	1		2.5	
Effluent SCOD	>35		15	

Table 17. Electrical Energy Requirement for Aeration, Aerobic Digestion and Sludge Recycle

October 88 to December 88	2.5 rotors used 1 Ditch in operation Aeration energy = 1930 KWH per day MCRT = 12 days
Phase V	
October 89 to December 89	Aeration energy = 2010 KWH per day
June 89 to May 90	Aeration energy = 1965 KWH per Day
RATIO (Fall to Annual)	2010/1965 = 1.023
Anaerobic Cell Pumps	Pumping energy = 330 KWH per day
Phase I	
Phase I (Oct 86 to Sep 87)	2.5 rotors used for secondary treatment 1 ditch used for secondary treatment MCRT = 8 to 12 days
Phase I (Oct 86 to Dec 86)	1930 KWH/day (assumed equal to Phase III)
Annual for Phase I	1930 /RATIO = 1930/1.023 = 1885 KWH per day
Aerobic Digestion energy	1.0 rotor operation (a conservative estimate) Energy required = 1/2.5 * 1885 = 750 KWH per day

	Phase I	Phase V
Aeration Energy	1885	1965
Aerobic Digestion Energy	750	
Anaerobic Cell Pumps		330
Total	**2635**	**2295**

Net Savings in Energy (Phase I - Phase V) = 2635 - 2295
= 340 KWH/day

= $7500 per year at $0.06 per KWH

During Phase V, an average of 330 KWH/day of was required for pumping mixed liquor from the anaerobic cell to ditch 1. This cost would have been lower if the anaerobic cell and the oxidation ditch were at the same elevation. At Bowie, the difference in elevation was 15 feet. Pumping costs were minimized by reducing the flow influent and recycle flow to the anaerobic cell to 30 to 60 percent of the average influent flow.

The total energy consumption for pumping, aeration and aerobic digestion decreased by 340 KWH per day from Phase I to Phase V. At a cost of 0.06 cents per KWH, this generated an annual savings of $7,500 (Table 17).

3) 3.57 mg/L of alkalinity is generated for each mg/L of nitrate-N denitrified. Several biological nutrient removal processes average 6 mg/L of nitrate-N in the effluent. Denitrification of nitrate-N from 6 mg/L to 2 mg/L at the Bowie WWTP generated an additional 14 mg/L of alkalinity. This reduced the annual expenditure on supplemental alkalinity added as caustic soda (30 percent at 70 cents per gallon) by $17,500 (Table 17).

The Bowie WWTP can operate at effluent alkalinity levels 10 mg/L less than the current operating range of 72 to 75 mg/L without violating the effluent pH permit. Since this buffering capacity is potentially available, it will reduce the cost for supplemental alkalinity by $12,500 annually (Table 17).

An annual net savings of $57,000 was observed in spite of the increase in nitrification which reduced the year round average for effluent ammonia-N from concentrations in excess of 10 mg/L in Phase I to 0.2 mg/L in Phase V. The effluent nitrate-N concentration was maintained at about the same level (between 1 and 3 mg/L). Annual net savings are computed in Table 18.

To operate the plant after new NPDES permits for nitrogen are implemented, laboratory technician time will have to be increased for nitrogen analysis. The level of training available to operators will also have to be increased. The maintenance procedures will have to be streamlined to increase the reliability of the aeration devices. The cost for additional laboratory facilities and technician time is estimated at $23,000 per year.

Table 18. Annual Savings in Operating Costs Generated by the BNR Process

ITEM	SAVINGS	
Direct (difference between 1987 and 1990) [2.2 MGD influent flow]		
Chemicals (Ferrous Sulfate)	$ 30,000	
Indirect [2.2 MGD influent flow]		
Caustic Soda for Chemical P Removal	$ 37,500	
Sludge Handling excluding lime stabilization	$ 0	
Aeration (from Table AERC)	$ 7,500	
Caustic Soda for Denitrification from 6 mg/L to 2 mg/L NO_3N	$ 17,500	
Buffer Available	$ -12,500	
Total		$ 80,000
Additional Lab Tech Time for Nitrogen monitoring (10 hours per week)	$ -17,000	
Laboratory Instruments (Depreciation) Chemicals for Nitrogen Monitoring	$ - 3,000	
Operator Training + Technical Assistance	$ - 3,000	
Net Savings		$ 57,000

A judicious investment in process control features, maintenance and operator training should reduce maintenance and operating costs in future. It will decrease problems with filamentous bacteria and increase process efficiency.

CONCLUSIONS

1. Biological nitrogen and phosphorus removals can be enhanced in oxidation ditch systems by controlling aeration to maintain reliable aerobic, anoxic and anaerobic volumes in the oxidation ditch(es).

2. The VT2-BNR process developed at Bowie uses a side stream anaerobic cell which receives between 40 and 60 percent of the influent flow. The anaerobic cell supplements biological phosphorus removal reactions which take place in the oxidation ditches.

3. The BNR process at Bowie is capable of maintaining less than 0.5 mg/L total phosphorus and between 3 and 4 mg/L for total nitrogen all year round. In its first year of operation, the effluent phosphorus averaged less than 0.7 mg/L (monthly average) in cold weather and 0.5 mg/L (monthly average) in warm weather. The annual average was 0.6 mg/L for total phosphorus without secondary effluent filtration.

The effluent ammonia-N averaged less than 0.5 mg/L (monthly average). The effluent TKN averaged 1 mg/L. The total nitrogen averaged 3.2 mg/L over a one year period of operation.

4. SVI can be controlled by installing process control features to continuously regulate the anaerobic, anoxic and aerobic volume fractions inside the oxidation ditches. Failure to control the volume fractions over a 12 to 36 hour period in warm weather can lead to significant increase in population of filamentous bacteria.

5. Several parameters determine the minimum operating D.O. set point of a BNR system at which operations may be optimized for BOD, nitrogen and

biological phosphorus removal. Additional control parameters such as alkalinity and turbidity may be used in a low rate, high nitrate recycle system.

6. A mathematical procedure was developed to determine an operating range for effluent alkalinity which would reflect the ammonia and oxidized nitrogen levels in the effluent.

Effluent Alkalinity = Available Influent Alkalinity - 3.57 (Influent TKN)
+ Effluent N forms effect

Effluent N forms effect = 3.57 (Eff Ammonia-N) - 3.57 (Eff Oxidized-N)

7. Aeration effluent turbidity can be used as a control parameter for nitrogen and phosphorus removal in a low rate system operating with a high nitrate recycle. The effectiveness of turbidity control increases with the magnitude of the nitrate recycle (internal recycle for oxidation ditches and counter current diffuser systems). Turbidity control may also be used in a high rate, low nitrate recycle system if turbidity is monitored in the first part of the aerobic zone.

8. Though the Bowie WWTP uses two oxidation ditches in series, a BNR process can be designed for a single loop oxidation ditch if it has an adequate circulation time and the capacity to maintain MCRTs which are adequate for nitrogen and phosphorus removal. Circulation time is the average length of time for the mixed liquor to flow (once) around the ditch. The minimum circulation time at which efficient treatment can be accomplished will depend on the resolution necessary on the control parameter to maintain aerobic and anoxic volumes.

9. There were significant reductions in operating costs for electrical energy, aeration, chemicals for phosphorus removal, and supplemental alkalinity. Supplemental alkalinity would have to be added if chemical P removal was practiced and biological denitrification was not as efficient. Total savings of $80,000 were observed for treating 2.2 MGD of flow (8300 m^3/day) containing an average of 175 mg/L of BOD_5, 450 mg/L of COD and 35

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APPENDIX

Appendix A

Key to Abbreviations for Monthly Average Tables

CFlow (MGD)	Combined Influent Flow
Recirc (MGD)	Plant Recycle from Solids Handling
FFlow (MGD)	Final Effluent Flow
RetF (MGD)	Return Sludge Flow
WASD (1000 GPD)	Waste Sludge Flow
AnaEn (KWH)	Energy consumed by Pumps in Anaerobic Basin per day (An energy expense related to bio-P removal)
RotEn (KWH)	Energy consumed per day in aeration
AtmT (F)	Atmospheric Temperature (Fahrenheit)
Tb (NTU)	Secondary Effluent Turbidity
TbMax	Maximum Secondary Effluent Turbidity (monthly average of maximum per day)
TbMin	Minimum Secondary Effluent Turbidity
DL1Ad (turns)	Adjustment to effluent weir of Ditch 1 (1 turn = 0.5 inch)
FDL2Ad (turns)	Adjustment to effluent weir of Ditch 2
NaOH (GPD)	Average NaOH consumption (25 percent)
SVI (mL/g)	Sludge Volume Index (measured in Ditch 2 in Phase V)
Dry Solids (lb/day)	Dry Solids Production (after dewatering)
AlkD1 (mg/L)	Alkalinity of mixed liquor supernatant in ditch 1
AlkD2 (mg/L)	Alkalinity of mixed liquor supernatant in ditch 2
SSD1 (mg/L)	Suspended Solids in ditch 1
SSD2 (mg/L)	Suspended Solids in ditch 2
AnaF1 (MGD)	Average Flow Rate to Anaerobic Zone
Note:	This flow is not continuous. It operates at this rate for 16 hours each day. The flow shuts off for 6 hours. To get 24 hour average, multiply flow by 0.66
MLRF1 (MGD)	Average Mixed Liquor Recycle (Continous flow over 24 hours)
WaterT (F)	Mixed Liquor Temperature (measured in ditch 2 in Phase V)

Individual Station Parameters

D.O. (mg/L)	Dissolved Oxygen
pH	pH
ORP (mv)	Oxidation Reduction Potential
T (C)	Liquid Temperature
Alk (mg/L)	Alkalinity
OP (mg/L)	Ortho-Phosphorus
TSP (mg/L)	Total Soluble Phosphorus
TP (mg/L)	Total Phosphorus
NH3N (mg/L)	Ammonia-N
SKN (mg/L)	Soluble Kjeldahl Nitrogen
TKN (mg/L)	Total Kjeldahl Nitrogen
NO2N (mg/L)	Nitrite-N
OxN (mg/L)	Oxidized-N (Nitrite-N + Nitrate-N)
SBOD (mg/L)	5 day soluble BOD

Individual Station Parameters, Continued

BOD (mg/L)	5 day total BOD
SCOD (mg/L)	Soluble COD
COD (mg/L)	Total COD
TSS (mg/L)	Total Suspended Solids
VSS (mg/L)	Volatile Suspended Solids
TS (mg/L)	Total Solids - Measured for thickened and dewatered sludge samples (instead of TSS)

Calculated Parameters

TN (mg/L)	Total Nitrogen (TKN + Ox-N)
MCRT (days)	Mean Cell Residence Time
% P, TSS basis	Percent Phosphorus in sludge, TSS basis
% P, VSS basis	Percent Phosphorus in sludge, VSS basis
% N, TSS basis	Percent Nitrogen in sludge, TSS basis
% N, VSS basis	Percent Nitrogen in sludge, VSS basis
DN (#)	Pounds of Nitrogen Denitrified per day
DN (mg/L)	mg/L of influent nitrogen denitrified (mg/L calculated in terms of influent flow)
% Vol Solids	Percent of Volatile Solids in Ditch 1 for Phase II and III, Ditch 2 for Phase V and average of two ditches for Phase IV.

Appendix B. Expanded Table of Monthly Averages

	CFlow MGD	Recirc	FFlow MGD	RetF MGD	WASF 1000GPD	AnaEn KWH	RotEn KWH	AtmT F	Tb NTU	TBMax	TbMin	DL1Ad turn	DL2Ad turn	NaOH GPD	SVI mL/g	Dry Solids lbs/day		AlkD1 mg/L	AlkD2 mg/L	AernAl	AlkAdj		SS m
Sep 87									3.6						190	3341		118		118	143		2
Oct 87									2.9						169	4349		107		107	132		3
Nov 87									3.5						143	3283		101		101	126		2
Dec 87									5.7						124	2259		112		112	137		2
Jan 88									6						121	3933		110		110	135		2
Feb 88	2.19								8						101	4515		113		113	138		2
Mar 88	2.21			1.68	37.5				8.6						103	4469		130		130	155		2
Apr 88	2.25			1.73	28				13.3						119	4765		123		123	148		2
May 88	2.37			1.79	27.2				4.7						112	4069		114		114	139		3
Jun 88	2.34			2	17.8				6.6						128	5021		71		71	96		3
Jul 88	2.55	0.45			42.1		2672	93.2							174	4440		71		71	96		3
Aug 88	2.56	0.52			27.8		2769	89.6							221	4983		71		71	83		3
Sep 88	2.54	0.46					2481	77.9							164	3776		94		94	94	94	2
Oct 88	2.15	0.11					2100	59	5.5						143	2409		106					2
Nov 88							2210	57.2	5.6						138	2897		119					3
Dec 88							1800	45	4.3						154	3442		96		96	96		3
Jan 89	2.66	0.15		2	25.7		1610	44				0.38			143	3590		95		95	95		
Feb 89	2.8	0.14		2			1843	42				0.75	1.25	28.6	165	4422		90	84				3
Mar 89		0.17	2.125	2	44.3		1958	52	3.5			2.11	2.48	33.3	142	5337		81	83				2
Apr 89	2.62	0.1	2.104	2	24.6	150	2117	63.2	2.9			1.23	1.37	7.4	131	4379		79	80				2
May 89	2.65	0.11	2.419	2	23.1	186	2167	70	2				0.63	39.5	114	4161		76	77				2
Jun 89	2.73	0.16	2.513	2	33.25	290	1600	89.6	1.9				1.43	2.4	134	5147		84	87				2
Jul 89	2.68	0.08	2.15	2	25.4	345	1844	86.7	1.7					0	147	4743		88	72	72	72		2
Aug 89	2.47		1.96	2		337	1982		1.3				3.88	0	151	2799		95	73	73	73		
Sep 89	2.42	0.055	1.97	2	30.1	0	1994	78.4	1.5				-1	0	177	3327		101	72	72	72		
Oct 89	2.08	0.066	1.95	2	27.8	323	2075	68.8	1.6				-3	0	185	2440		97	72	72	72		
Nov 89	1.97	0.06	1.86	2	25.5	321	1965	50.2	2				-1	0	155	3459		97	73	73	73		
Dec 89	2.21	0.066	1.96	2	28.1	317.1	1991	30.6	1.6				-1.5	0	155	3137		99	75	75	75		
Jan 90	2.42	0.067	2.06	3	26	308	1983	47.9	1.5				1.25	0	148	3030		93	75	75	75		
Feb 90		0.055	1.99	3	25.5	302	1890	51	1.9				-2.25	0	124	3973		94	75	75	75		
Mar 90	2.05	0.059	1.97	3	29.1	336	2076	54.2	2.1					0	128	3975		98	73	73	73		
Apr 90		0.068	2.07	2		346	1963		2.1					0	116	4364		101	77	77	77		
May 90			2.11	2		380	2200								119								
Jun 90						390	2236																
Ph II	2.318	0.45	ERR	1.8	30.52	ERR	2672	93.2	6.589	ERR	ERR	ERR	ERR	ERR	129.4	4110	ERR	105.2	ERR	105.2	130.2	ERR	3
Ph III	2.45	0.24	ERR	2	25.7	ERR	1930	56.62	5.133	ERR	ERR	0.38	ERR	ERR	148.4	3223	ERR	102	ERR	95	95	94	3
Ph IV	2.69	0.13	2.216	2	30.67	168	2021	56.8	2.8	ERR	ERR	1.363	1.433	27.2	138	4575	ERR	81.5	81	ERR	ERR	ERR	2
Ph V	2.337	0.074	2.047	2.25	27.86	300.4	1964	61.93	1.745	ERR	ERR	ERR	-0.27	0.218	144.9	3672	ERR	95.18	74.91	73.7	73.7	ERR	2

	SSD2		AnaFl	MLRFl	WaterT	BW05 Raw Influent					BW10 Comb Influent										BW15 Anaerobic C		
	mg/L		MGD	MGD	F	TP mg/L	TKN mg/L	COD mg/L	BOD mg/L	TSS mg/L	DO mg/L	pH	T C	Alk mg/L	OP mg/L	TP mg/L	NH3 mg/L	TKN mg/L	SCOD mg/L	COD mg/L	D.O. mg/L	ORP mv	P
Sep 87																8.55	17.7						
Oct 87																9.26	25.6						
Nov 87																9.35	24.7			459			
Dec 87																7.12	28.9			403			
Jan 88																6.1							
Feb 88														150	1.94	8.21	27.6	47		488			
Mar 88														171	1.93	8.29	28.4	49.6		487			
Apr 88					57									176	1.37	7.72	23.9	42.6		457			
May 88					60									182	1.93	6.94	25.4	40.2		415			
Jun 88					60									172	2.27	7.22	28.9	45.6		453			
Jul 88					66						1.03	7.34	23	180	1.9	6.09	21.7	32.8	126	363			
Aug 88					75						0.66	7.33	24.2	170	1.99	12.5	18.5	42.1	109	742			
Sep 88					78						0.84	7.14	22.8	176	1.74	7.17	18.5	28.9	89	373			
Oct 88					74						1.3	7.24	21.4	184	2.41	7.8	22.1	36.6	117	460			
Nov 88					69											8.27		38					
Dec 88					66											6.83		34.4					
Jan 89					61	8.58	48.3	833	230	250	3.2	7.5	15.5	155	3.28	8.14	26.6	43.2	167				
Feb 89	3072				58	6.81	39.7	484	185	214	4.35	7.5	14.8	154	2.3	9.78	20.9	42.7	179				
Mar 89	2741				57	6.82	38.1	475	195	223	4.64	7.7	15	168	2.5	10.78	20	40.5	196				
Apr 89	2441		0.4	1	56	7.44	40.2	523	220	251	4	7.7	16.3	178	2.16	8.4	20.5	38	225		0.01		
May 89	2576		0.6	1	62	6.17	34.1	432	163	187	3.2	7.5	17.8	160	2.06	7.05	18.4	32.9	183		0.03		
Jun 89	2996		1.8	1.5	64	5.73	31.8	406	179	235	2.2	7.3	21.2	139	1.36	5.61	17	26.5	161		0.07		
Jul 89	2666		1.8	1.44	73	6.84	36.3		107	196	1.3	7.4	22	146	2.71	9.09	17.9	36.8			0		7.
Aug 89	2503		1.8	1.44	74	6.4																	
Sep 89	3032		0	0	76	7.48	35.5	450	167	200	2.37	7.5	23	156	3.7	8.46	18.8	31.7	192				
Oct 89	2940		1.76	1.21	72	7.86	43.2	420	147	184	1.4	7.5	21.7	167	3.82	9.15	18.6	39.3	146		0.01	-72	
Nov 89	3328		1.73	1.11	69	6.6	37.3	519	172	233	1.72	7.5	19.6	171	3.8	11.2	17.8	38.3	244		0	-39	
Dec 89	3690		1.53	1.07	63	7.28	38.5	506	158	226	3.53	7.6	15.8	190	4.8	13.8	24.9	42.9			0.025	24	
Jan 90	4042		1.34	1.07	54	6.18	35.5	450	174	219	5.26	7.6	15	183	3.3	10.1	25.2	39.8			0.02	-27	
Feb 90	3950		1.35	1.08	56	5.73	29.7	361	160	222											0	-75	
Mar 90	4113		1.38	1.07	57	5.33			178	217											0.1	4	
Apr 90	3427		1.39	1.19	60	5.3	44.2																
May 90																							
Jun 90																							
Ph II	ERR	ERR	ERR	ERR	60.75	ERR	ERR	ERR	ERR	ERR	1.03	7.34	23	171.8	1.89	7.63	26.12	42.97	126	440.6	ERR	ERR	ER
Ph III	ERR	ERR	ERR	ERR	69.6	8.58	48.3	833	230	250	1.78	7.293	19.9	171.7	2.477	7.642	22.4	36.22	124.3	416.5	ERR	ERR	ER
Ph IV	2708	ERR	0.5	1	58.25	6.81	38.03	478.5	190.8	218.8	4.048	7.6	15.98	165	2.255	9.003	19.95	38.53	195.8	ERR	0.02	ERR	6.
Ph V	3335	ERR	1.444	1.107	65.27	6.43	36.89	444.6	160.2	214.7	2.54	7.486	19.76	164.6	3.356	9.63	20.03	36.47	185.8	ERR	0.028	-30.8	7.1

	Cell											Before Aerator 1-2													
	OP mg/L	NH3N mg/L	NO2N mg/L	OrN mg/L	SCOD mg/L	COD mg/L	TSS mg/L	VSS mg/L	BW25 DO mg/L	Ditch 1 ORP mv	OP mg/L	NH3N mg/L	OrN mg/L	SCOD mg/L	TSS mg/L	VSS mg/L	BW30 DO mg/L	Ditch 1 ORP mv	pH	T C	OP mg/L	TP mg/L	N m
Sep 87																							
Oct 87																							
Nov 87																							
Dec 87																							
Jan 88																							
Feb 88																							
Mar 88										112					2918	2200	0.32	112	7.1	13.7	0.2	74.2	
Apr 88									0.42	122	0.27		0.02		2560	1936	0.55	122	6.9	15.6	0.13	65.6	
May 88									1.42	168	0.6		1.05		3209	2265	1.53			19			
Jun 88									0.94	191	0.41	0.27	2.83				0.62			20.9	0.36		
Jul 88									0.25		0.06	1.9	1.04		3726	2665	0.22		6.8	24.4	0.05	104	
Aug 88									0.22		0.42	1.58	1.06		3099	2409	0.27		6.8	25.7	0.37	90.6	
Sep 88									0.25		0.86	4.64	0.08		2828	2210	0.22		6.7	23.5	0.77	72	
Oct 88									0.23		0.25	1.99	0.72		2637	2080	0.25		6.8	21	0.22	93	
Nov 88																							
Dec 88																							
Jan 89									1.55		0.07	10.4	0.16		3800	3200	1.4			14.3	0.05	108	
Feb 89																	0.3		6.9	13.75	0.72	119	
Mar 89																	0.46		6.91	13.3	0.64	94	
Apr 89	3.65	4.24	0.02	0.06	24.4	2732	1950	1507									0.26		6.85	15.9	0.71	73	
May 89	4.26	7.41	0.02	0.07	38.1	2027	1447	1132									0.3		6.91	18.3	0.8	71	
Jun 89	3.51	16.35	0.01	0.14	24	1824	1390	1095									0.03		6.93	22.5	0.18	83	
Jul 89	5.14	15.37	0.01	0.08	30.8		1340	1086									0		6.94	23.6	0.7	89	
Aug 89					58																		
Sep 89									0	115	1.87	8.63	0.09		3302	2835			7.1	23.1	1.64	127	
Oct 89	7.75	17.5	0.01	0.03	55.9	1677	1434	1136									0.01		7.1	21.4	1.59	108	
Nov 89	6.97	16.8	0.03	0.64	58.9	2030	1916	1523									0.01		7	18.5	2.1	132	
Dec 89	8.06	20.1	0.01	0.02	40.7	2411	2015	1615									0.05	54	7	13.8	1.68	137	
Jan 90	8.93	17.6	0.01	0.04	48.6	2622	2217										0.12		7	14	0.82	152	
Feb 90				0.03	45.7		2060	1680	0	-32.3	0.54	0.05	0.5		3910								
Mar 90																							
Apr 90																							
May 90																							
Jun 90																							
Ph II	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	0.758	148.3	0.335	1.085	1.235	ERR	3103	2267	0.648	117	6.933	18.72	0.185	81.27	1
Ph III	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	0.677	ERR	0.393	5.677	0.32	ERR	3088	2497	0.623	ERR	6.75	19.6	0.347	91	4
Ph IV	3.955	5.825	0.02	0.065	31.25	2380	1699	1320	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	0.33	ERR	6.893	15.31	0.718	89.25	0
Ph V	6.727	17.29	0.013	0.14	45.33	2113	1767	1356	0	41.35	1.205	4.34	0.295	ERR	3606	2835	0.037	54	7.01	19.56	1.244	118.3	5

	TKN mg/L	NO2 mg/L	OxN mg/L	SCOD mg/L	COD mg/L	TSS mg/L	VSS mg/L	BW31 DO mg/L	Ditch 2 ORP mv	pH	T C	OP mg/L	TP mg/L	NH3 mg/L	TKN mg/L	NO2 mg/L	OxN mg/L	SCOD mg/L	COD mg/l	TSS mg/L	VSS mg/l	BW35 OP	NI
Sep 87																							
Oct 87																							
Nov 87						2676	1875																
Dec 87						2008	1547																
Jan 88																							
Feb 88																							
Mar 88	205	0.01	0.01			2918	2200																
Apr 88	176	0.01	0.04			2517	1869																
May 88																							
Jun 88			3.25																				
Jul 88	216	0.29	1.32	23.9	3866	3726	2665																
Aug 88	180		1.14	28.7	2988	3099	2408																
Sep 88	153	0.1	0.14	43.5	3173	2574																	
Oct 88	202	0.46	0.72	39.4	4219	2923																	
Nov 88																							
Dec 88																							
Jan 89	325	0.04	1.24	33.4	5639	4120																	
Feb 89	211	0.19	0.87	19.05	4102	3648	2763	0.45		6.93	13.8	0.74		1.51			0.92	19.4		3170	2348		
Mar 89	195	0.26	1.8	24.2	3665	2784	2120	0.46		6.9	13.3	0.79		1.19		0.23	1.22	29.8		2845	2172		
Apr 89	152	0.61	1.72	25.9	2933	2215	1705	0.26		6.89	15.9	0.56	82.6	0.19	167		0.96	20.5		2510	1912		
May 89	145	0.1	1.88	20.7	2863	2085	1587	0.3		6.93	18.3	0.67	82.6	0.65	163	0.04	1.22	20.4		2455	1880		
Jun 89	173	0.05	0.72	16.5	3192	2590	1985	0.05		6.87	23.2	0.1	89.3	0.53	183	0.03	1.26	11.95		2740	2085		
Jul 89	174	0.09	0.4	18		2505	1933	0.02		6.91	23.8	0.2	87.4	0.42	169	0.07	1.44	20.4		2546	2010		
Aug 89				32														17					
Sep 89	214	0.01	0.08	21.6	3772	2920	2207	0.16	171	6.98	22.8	0.66	114	0.48	197	0.07	4.16	17.9		2960	2222		
Oct 89	188		0.32	53.6	3457	2856	2178	0.8	40	7.1	21.3	0.58	121	0.08	211	0.03	3.49	30.2		2816	2136		
Nov 89	228		0.37	34.9		3703	2860	0.83	82	6.95	18.1	0.5	130	0.23	216	0.07	3.86	18.4		3453	2620		
Dec 89	254		0.68	22.6		3608	2825	0.84	116	7	13.4	0.45	147	1.54	265	0.04	3.61	19.7		3805	2960		
Jan 90	294		1.05	20.8		4205		0.54	69	6.9	13.7	0.3	153	0.03	283	0.02	2.32	20.2		4173	3295		
Feb 90								0.34	25	6.8			145	0.04	300			21.4					
Mar 90								0.6	35	7													
Apr 90																							
May 90																							
Jun 90																							
Ph II	199	0.103	1.155	23.9	3866	2769	2031	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	EI
Ph III	226.7	0.2	0.7	38.77	4344	3206	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	EI
Ph IV	175.8	0.29	1.568	22.46	3391	2683	2044	0.368	ERR	6.913	15.33	0.69	82.6	0.885	165	0.135	1.08	22.53	ERR	2745	2078	ERR	EI
Ph V	217.9	0.05	0.517	27.5	3474	3198	2331	0.464	76.86	6.946	19.47	0.399	123.3	0.419	228	0.047	2.877	19.68	ERR	3213	2475	ERR	EI

	OxN	SCOD	TSS	BW40 OP	NH3	OXN	SCOD	TSS	VSS	BW55 OP	NH3	OxN	SCOD	TSS	BW62 OP	BW65 OP	TP	NH3	TKN	OxN	COD	TSS	
Sep 87																							
Oct 87																							
Nov 87																							
Dec 87																							
Jan 88																							
Feb 88																							
Mar 88																							
Apr 88																							
May 88																							
Jun 88																							
Jul 88																							
Aug 88																							
Sep 88																							
Oct 88																							
Nov 88																							
Dec 88																							
Jan 89																							
Feb 89																							
Mar 89										7.23	3.74	0.03		4270									
Apr 89										7.94	1.74	0.09		1721		0.12	31.83	25.6	103.3	0.16	1314	890	
May 89										8.12	5.09	0.04		125		0.11	18.9	47.7	85.3	0.34	1108	540	
Jun 89															1.28								
Jul 89										5.02	1.07	0.06			0.97								
Aug 89																							
Sep 89										4.88	1.4	0.31	25.35										
Oct 89										20.7	2.94	0.11	49.95										
Nov 89										13.62	2.17	0.02	38.8										
Dec 89										22.2	5.1	0.06	59										
Jan 90										3.99	1.34	0.01	28.6										
Feb 90										5.5	2.24	0.07	38										
Mar 90																							
Apr 90																							
May 90																							
Jun 90																							
Ph II	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	El
Ph III	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	El
Ph IV	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	7.763	3.523	0.053	ERR	2039	ERR	0.115	25.37	36.65	94.3	0.25	1211	715	El
Ph V	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	ERR	10.84	2.323	0.091	39.95	ERR	1.125	ERR	ERR	ERR	ERR	ERR	ERR	ERR	El

	BW70 OP mg/L	TSP mg/L	Final Effluent TP mg/L	NH3 mg/L	SKN mg/L	TKN mg/L		NO2 mg/L	OxN mg/L	SCOD mg/L	COD mg/L	BOD mg/L	TSS mg/L	VSS mg/L	DO mg/L	pH	T C	Alk mg/L	TN mg/L	Calculated Parameters MCRT days	%P TSS basis	%P VSS	
Sep 87	0.24		0.45	13.5		16.5	3		0		26	7							16.5	14.9			
Oct 87	0.22		0.47	9.83		12.8	2.97		0.2		29	26	8						13	10.8			
Nov 87	0.32		0.66	7.62		10.4	2.78		0.4		35.3	9.5	8						10.8	13.9			
Dec 87	0.5		0.8	17		20.8	3.8		0		65	20.3	22						20.8	14.9			
Jan 88			0.9	18		21	3		0.01		66	18	31						21.01	9.6			
Feb 88	0.34		1	19.5		21.6	2.1		0.01		69.3	16.8	28				12.8	141	21.61	10.2			
Mar 88	0.4		0.84	20.1		24.5	4.4	0.01	0.01		84.5	25	16				13.5	152	24.51	10.2	2.52		
Apr 88	0.48		0.77	18.8		23.1	4.3	0.01	0.02		94.5	19.2	28				15.8	147	23.12	7.73	2.49		
May 88	0.52		0.87	17.2		20.1	2.9	0.32	0.44		47.5	9.2	13				19.2	131	20.54	10.6	2.98		
Jun 88	0.53		0.79	0.95		2.56	1.61	0.52	0.68		34	6	18				21.4	79	3.24	10.9	3.17		
Jul 88	0.34		0.67	0.51		1.83	1.32		1.72	33	42.3	14.6	10				24.9	69.8	3.55	11.5	2.79		6
Aug 88	1.31		1.83	0.56		2.17	1.61		1.31	28.8	38	13.8	8				25.7	72	3.48	8.6	2.92		5
Sep 88	1.91		2.24	2.69		4	1.31		3.01	36.9	48.7	22	5				23.4	101	7.01	10.3	2.81	3.57	5
Oct 88	1.53		1.79	1.21		2.71	1.5		1.82	30.1	35.1	16	4				20.7	99	4.53	16.6	3.17	4.02	6
Nov 88			1.51	0.18		1.6	1.42		1.77			27	22						3.37	12			
Dec 88			1.99	5.84		6.92	1.08		3.11			20	20						10.03	13			
Jan 89	1.61		2.25	4.99		6.52	1.53	0.17	3.77	36.4	53	9	12				14	125	10.29	13			
Feb 89	1.66		1.95	0.44		1.6	1.16	0.07	1.68	22.23	25.2	12	11				13	81	3.28	15		4.41	
Mar 89	1.55		1.69	0.52		1.22	0.7	0.07	1.81	27.9	30.7	7	9.8				13	80	3.03	14.2		4.42	
Apr 89	0.89		1.2	0.17		1.51	1.34		0.79	19.8	27.5	4	7.1					83	2.3	16.5		4.3	
May 89	0.81		0.88	0.28		0.94	0.66		1.89	20.5	25	5	4.9					82	2.83	18		4.48	
Jun 89	0.32		0.59	0.13		0.99	0.86		1.18	13.8	22	6	2.9				23.3	73	2.17	18	3.92	4.28	
Jul 89	0.35		0.43	0.1		0.8	0.7		1.27	9.95	30.2	4	0.6				23.9	73	2.07	17	4.15	4.35	
Aug 89			0.5	0.1		0.8	0.7		1.5			5	2						2.3	24.6			
Sep 89	0.79		0.97	0.1		0.68	0.58		2.98	12.4	25.6	3	3				23	80	3.66	27.6	3.85	5.14	6
Oct 89	0.71		0.9	0.06		0.82	0.76		3.55	19.85	29.7	3	6				20.8	75	4.37	36.5	4.33	5.7	7
Nov 89	0.54		0.63	0.18		0.93	0.75		3.9	14.6	18.4	3	7				17.7	72	4.83	29.1	3.76	4.96	6
Dec 89	0.47		0.57	1.1		1.2	0.1		2.4	17.44	19	3	6				13	80	3.6	35.6	3.86	4.97	6
Jan 90	0.61		0.64	0.1		0.92	0.82		2.77	17.5	18.4	3	3				13.5		3.69	40.3	3.73	4.66	7
Feb 90			0.55			1.1	1.1					3	3							30.1	3.99	4.98	8
Mar 90			0.78			0.7	0.7		3.5			3	3						4.2	31.2			
Apr 90			0.48			1.1	1.1		1.7			3	2						2.8	25.2			
May 90			0.35			1	1		1.5			3	2						2.5	21.1			
Jun 90							0																
Ph II	0.406	ERR	0.777	12.95	ERR	15.87	2.918	0.215	0.349	33	56.74	16.46	18.2	ERR	ERR	ERR	17.93	120	16.22	11.03	2.79	ERR	6.
Ph III	1.683	ERR	1.956	2.982	ERR	4.35	1.368	0.17	2.696	34.47	45.6	18.8	12.6	ERR	ERR	ERR	19.37	108.3	7.046	12.98	2.99	3.795	6.
Ph IV	1.228	ERR	1.43	0.353	ERR	1.318	0.965	0.07	1.543	22.61	27.1	7	8.2	ERR	ERR	ERR	13	81.5	2.86	15.93	ERR	4.403	EI
Ph V	0.541	ERR	0.616	0.234	ERR	0.92	0.764	ERR	2.386	15.08	23.33	3.5	3.375	ERR	ERR	ERR	19.31	75.5	3.29	28.03	3.949	4.88	7.1

	%N VSS	DN mg/L Inf flow tests	%Vol Solids
Sep 87			
Oct 87			
Nov 87			
Dec 87			
Jan 88			
Feb 88			
Mar 88	8.24		75.4
Apr 88	7.98		75.6
May 88	9.05		70.5
Jun 88	8.09		
Jul 88			71.5
Aug 88			77.7
Sep 88	7.56		78.8
Oct 88	8.77		78.9
Nov 88			
Dec 88			
Jan 89			
Feb 89			74
Mar 89			76
Apr 89			77
May 89			76
Jun 89			76
Jul 89			79
Aug 89			
Sep 89	8.87	18.2	75
Oct 89	9.91	27.6	75.6
Nov 89	8.26	19.7	75.9
Dec 89	8.97	21.2	77.8
Jan 90	8.68	19.6	79
Feb 90	10.3	5.1	80
Mar 90			
Apr 90			
May 90			
Jun 90			
Ph II	8.34	ERR	73.25
Ph III	8.165	ERR	78.85
Ph IV	ERR	ERR	75.75
Ph V	9.165	18.57	77.29

AUTHORS

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