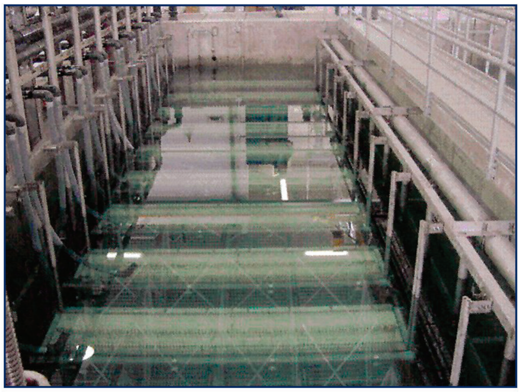
Municipal Nutrient Removal Technologies Reference Document

Volume 2 – Appendices



U.S. Environmental Protection Agency
Office of Wastewater Management, Municipal Support Division
Municipal Technology Branch

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Appendix A: Case Studies

Appendix A provides detailed case studies with information from nine wastewater treatment facilities selected for their excellent performance and varying technologies. Two facilities were chosen because of their denitrification technologies, two were chosen because of their phosphorus removal technologies, and an additional five facilities were included because of both nitrogen and phosphorus removal technologies.

Denitrification

- Central Johnston County, North Carolina
- Lee County, Florida

Phosphorus removal

- Kalispell, Montana (biological phosphorus)
- Clark County, Nevada (biological phosphorus and chemical phosphorus)

Nitrogen and phosphorus removal

- Kelowna, British Columbia (biological nitrogen and phosphorus)
- Marshall Street in Clearwater, Florida (biological N and chemical phosphorus)
- Noman Cole in Fairfax County, Virginia (biological nitrogen and chemical phosphorus)
- North Cary, North Carolina (biological nitrogen and phosphorus)
- Western Branch in Upper Marlboro, Maryland (three separate activated-sludge systems operated in series)

Acknowledgements

EPA and the authors would like to acknowledge the commitment, ingenuity and leadership demonstrated by the owners and personnel at the plants represented by the data reported in this document. The case studies in this document represent significant accomplishments made by the leaders of the facilities and their dedicated personnel. EPA recognizes their cooperation and assistance in providing information on their facilities. Permit compliance was achieved under all conditions, even under tropical storm conditions in North Carolina and under an extreme heat wave in Nevada. Some plants (Fairfax County, Virginia, and Clark County, Nevada), recognized as environmental leaders in their regions, are providing levels of treatment that go beyond their permit requirements. Central Johnston County, North Carolina, retrofitted an existing aeration system for biological phosphorus removal and nitrogen removal and developed the denitrification sludge blanket; Kelowna, British Columbia, and Lee County, Florida, made similar modifications. Kalispell, Montana, developed ways to minimize recycle loads from its sludge-handling processes while producing the lowest phosphorus concentration achieved entirely by a biological process. Clark County, Nevada, has a *Process Today's Sludge Today* policy. Clearwater, Florida, developed a control strategy for nitrogen removal on the basis of three sensors, producing a low nitrogen concentration in the effluent. Kalispell, Montana, is a good example of sound technical analysis carried out daily by the plant personnel in optimizing the phosphorus removal with the best reliability.

Central Johnston County Wastewater Treatment Plant
Smithfield, North Carolina
Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

The Central Johnston County Wastewater Treatment Plant (WWTP) is in Smithfield, North Carolina. The facility is designed for a capacity of 7 million gallons per day (MGD), and it processed an average of 4.12 MGD during the evaluation period, October 2005 to September 2006.

The plant was selected as a case study because it achieves a high level of biological nitrogen and phosphorus removal through a unique plug-flow, activated-sludge (AS) process retrofitted to the existing facility, followed by a new stand-alone denitrification filter process. The relevant National Pollutant Discharge Elimination System (NPDES) permit limits for the facility are shown in Table 1.

Table 1. NPDES permit limits

Parameter	Annual loading (lb)	Quarterly (mg/L)	Monthly average (mg/L)	Weekly average (mg/L)
BOD_5, 4/1–10/31			5	7.5
BOD_5, 11/1–3/31			10	15
TSS			30	45
Ammonia-Nitrogen, 4/1–10/31			2	6
Ammonia-N 11/1–3/31			4	12
Total phosphorus		2	1	--
Total nitrogen	56,200[a]			

Notes:
BOD_5 = biochemical oxygen demand
mg/L = milligrams per liter
P = phosphorus
TSS = total suspended solids
[a] Equivalent to 3.7 mg/L at 5 MGD

Plant Process

The plant layout is shown in Figure 1, and the process schematic is shown in Figure 2. After bar screens, wastewater flows first to anoxic basin 5, then to aerobic basin 4 or 6. The flow then goes to aerobic basin 1, 2, or 3 before secondary clarification and going through the denitrifying filters. Following ultraviolet disinfection, the water is discharged to the Neuse River. Biosolids are aerobically digested, dewatered, and hauled to a landfill.

Basis of Design and Actual Flow

Flow

The design flow for the facility is 7 MGD. The average flow for the study period was 4.12 MGD, while the maximum month flow during the study period was 5.17 MGD during June 2006. The maximum month flow occurred when Tropical Storm Alberto subjected North Carolina to very heavy rains.

Loadings

Plant loadings were as follows:

Anoxic basin 5: 1 million gallons (MG), or 4.8 hours
Aerobic basin–large: 1 MG, or 4.8 hours
Aerobic basin–small, 1 and 2: 0.55 MG, or 1.9 hours
Aerobic basin–small, 3: 0.34 MG, or 1.2 hours
Total hydraulic retention time (HRT): 11.5 hours
Internal recirculation rate: 8,000–12,000 gallons per minute (gpm), or four times the influent flow rate
Secondary clarifier: 6.7 hours, or 412 gallons per day per square foot (gpd/ft^2)
Denitrification filter, hydraulic loading rate: 3 gpm/ft^2

Plant influent and effluent average results for the period October 2005 to September 2006 are shown in Table 3.

Table 4 presents plant monthly averages for process parameters.

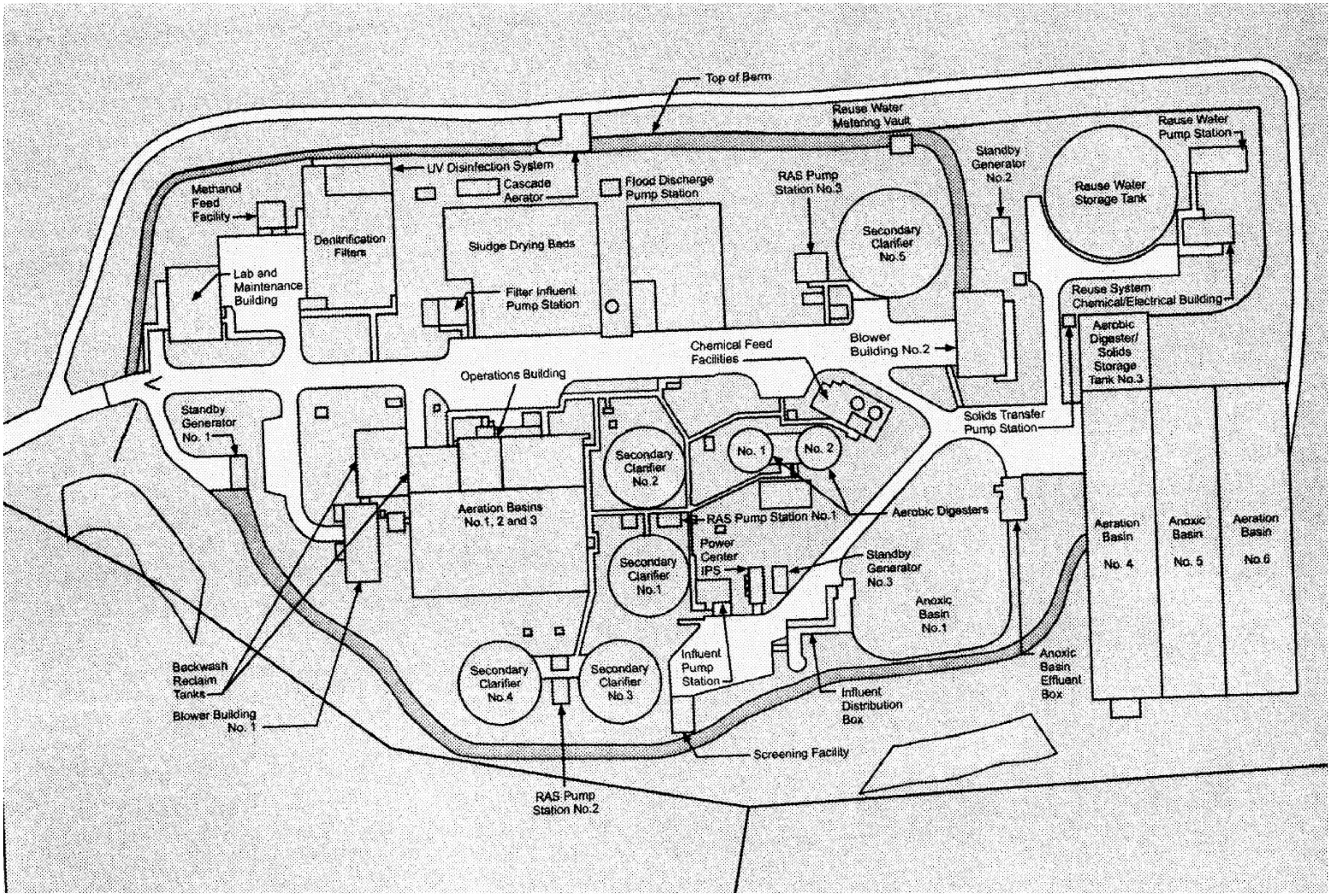

Figure 1. Central Johnston County WWTP layout.

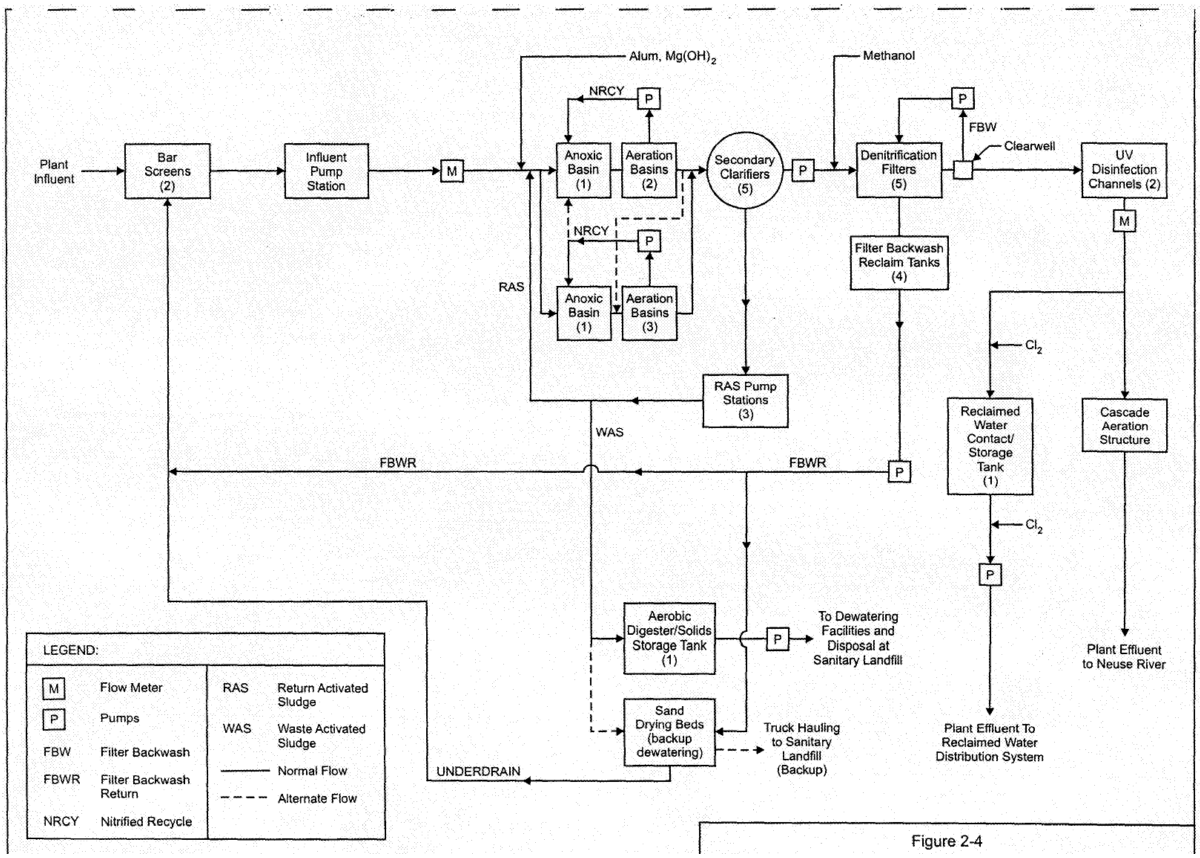

Figure 2. Central Johnston County WWTP process schematic.

Table 3. Influent and effluent averages

Parameter	Average	Max month	Max month vs. avg.	Max week	Sample method/frequency
Flow (MGD)	4.12	5.17	25%	6.2	--
Influent TP (mg/L)	5.8	8.5	46%	13.6	Weekly/composite
Effluent TP (mg/L)	0.26	0.64	140%	1.01	Weekly/composite
Influent BOD (mg/L)	320	386	20%	497	Daily/composite
Effluent BOD (mg/L)	3	4.59	32%	5.2	Daily/composite
Influent TSS (mg/L)	328	419	27%	564	Daily/composite
Effluent TSS (mg/L)	1.21	1.47	13%	1.8	Daily/composite
Influent NH_4-N (mg/L)	28	34.4	27%	37.4	Daily/composite
Effluent NH_4-N (mg/L)	0.44	0.54	22%	0.86	Daily/composite
Influent TN (mg/L)	31.2	42.7	37%	63.1	Daily/composite
Effluent TN (mg/L)	2.14	2.77	30%	3.13	Daily/composite

Notes:
BOD = biochemical oxygen demand
mg/L = milligrams per liter
NH_4-N = ammonia measured as nitrogen
TN = total nitrogen
TP = total phosphorus
TSS = total suspended solids

Table 4. Monthly averages for plant process parameters

Month	MLSS (mg/L)	Sludge age (d)	HRT (hr)	Temperature (°C)
Oct 2005	2,527	8.1	23.1	23
Nov 2005	2,445	7.9	13.9	19
Dec 2005	2,650	8.5	15.9	17
Jan 2006	2,686	8.6	15.1	16
Feb 2006	2,452	7.9	16.4	14
Mar 2006	2,643	8.5	23.6	16
Apr 2006	2,679	8.6	27.9	18
May 2006	2,417	7.8	23.5	20
June 2006	2,300	7.4	19.4	24
July 2006	2,378	7.6	23	26
Aug 2006	2,448	7.9	25.1	27
Sep 2006	2,574	8.3	21.6	25

Notes:
HRT = hydraulic retention time
MLSS = mixed liquor suspended solids

Performance Data

Figures 3 and 4 present reliability data for removal of total phosphorus (TP). The removal is good, with the effluent TP averaging 0.26 mg/L and a medium coefficient of variation (COV) of 62 percent.

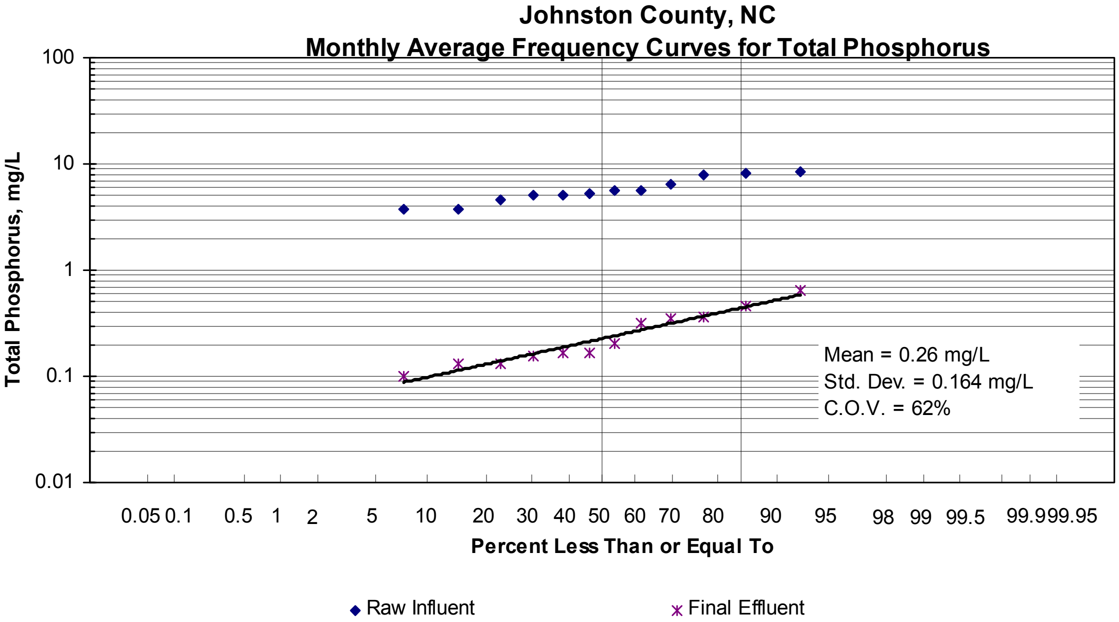

Figure 3. Monthly average frequency curves for TP.

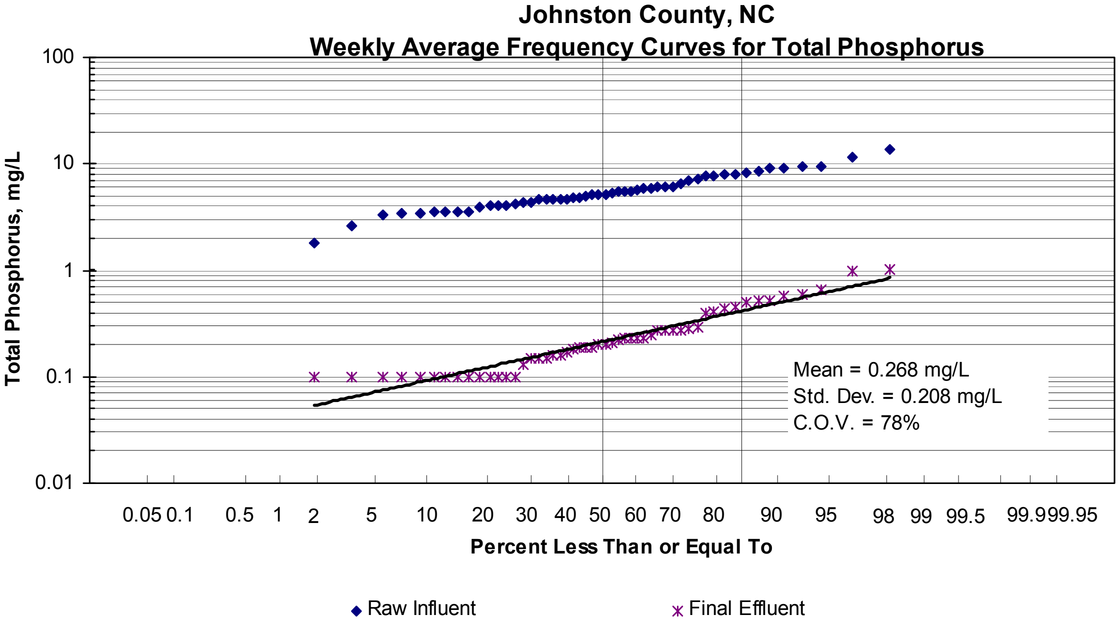

Figure 4. Weekly average frequency curves for TP.

Figures 5 and 6 present reliability data for ammonia nitrogen removal. Removal of ammonia nitrogen is very good, with a mean effluent of 0.44 mg/L and a very low COV of 12 percent.

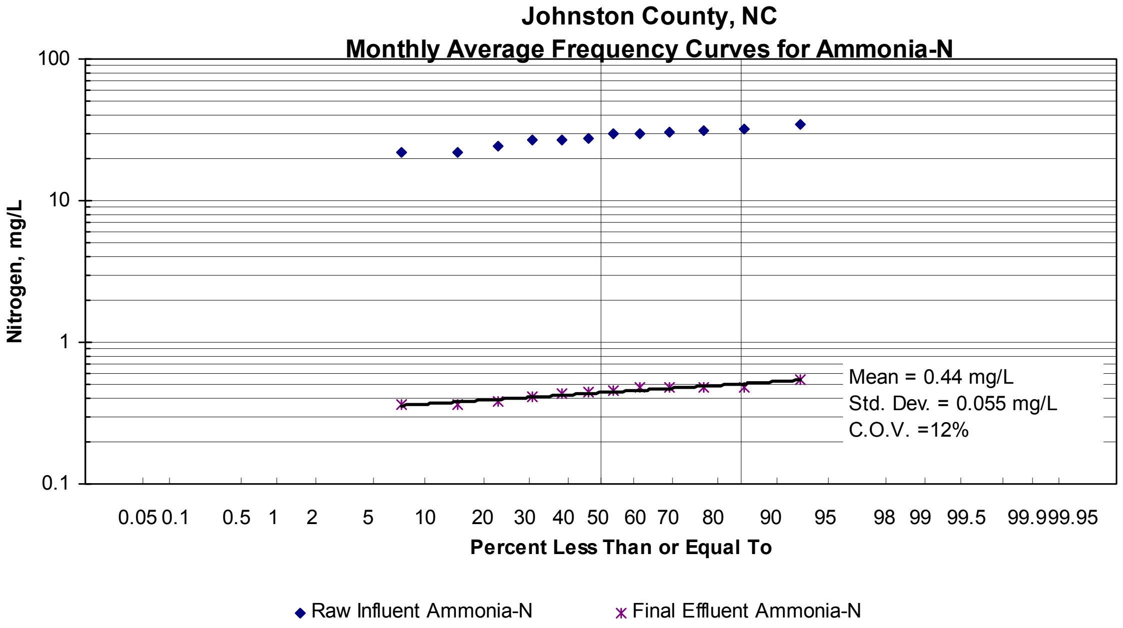

Figure 5. Monthly average frequency curves for ammonia nitrogen.

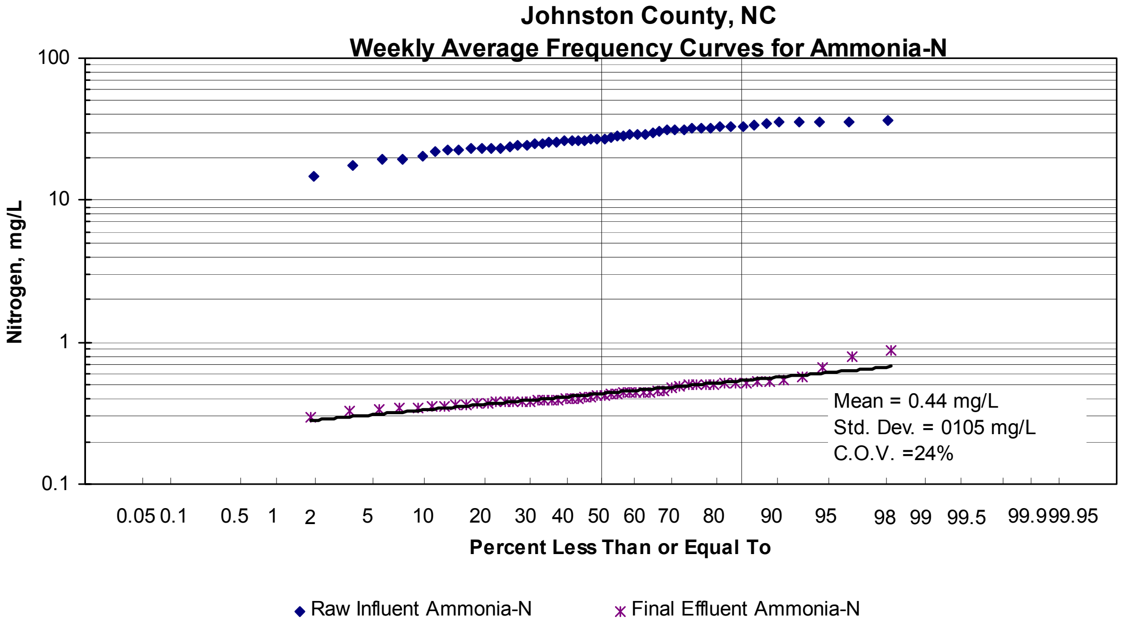

Figure 6. Weekly average frequency curves for ammonia nitrogen.

Figures 7 and 8 present reliability data for removal of total nitrogen (TN). Between the anoxic portion of the AS system and the denitrification filter, the plant gives outstanding TN removal, with effluent TN of 2.14 mg/L and a COV of 1 percent.

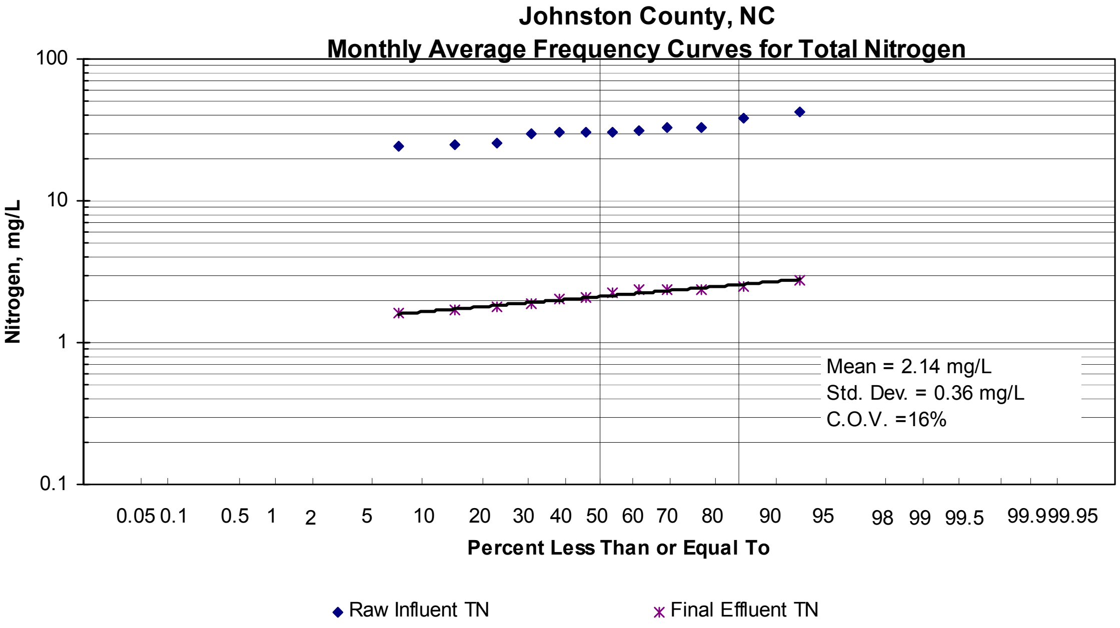

Figure 7. Monthly average frequency curves for TN.

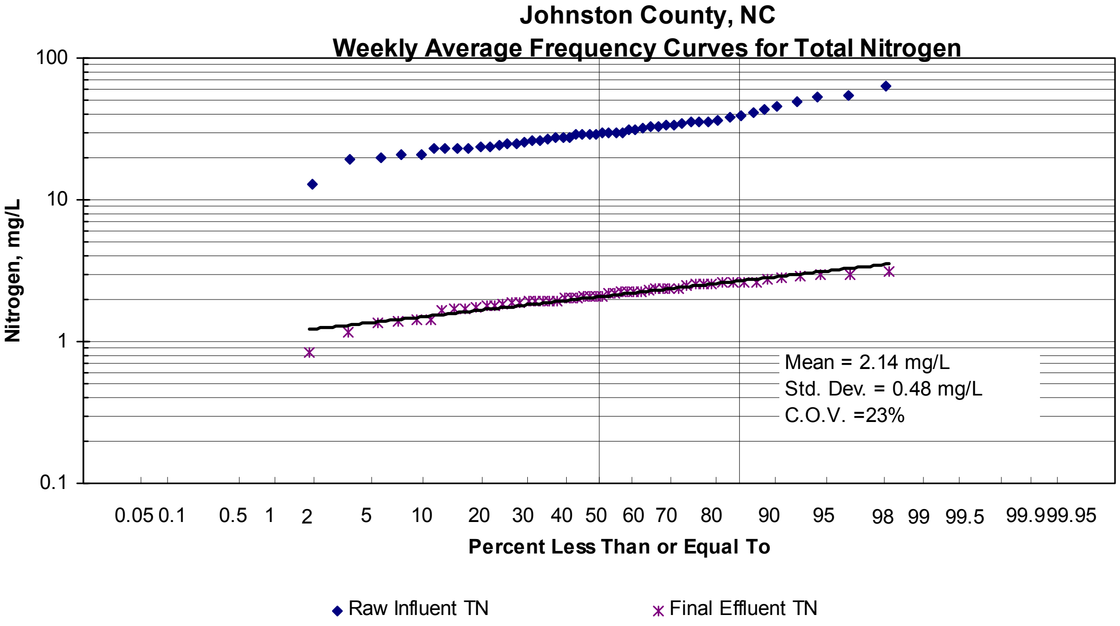

Figure 8. Weekly average frequency curve for TN.

Reliability Factors

This facility is unique in two areas: (1) biological phosphorus removal and nitrogen removal in a plug-flow, AS process and (2) separate-stage denitrification filters. The results are excellent. The plant achieves a phosphorus mean concentration of 0.26 mg/L with a COV of 62 percent without any chemical addition and a TN concentration of 2.14 mg/L with a COV of only 16 percent. The key factors for this exceptional performance are briefly discussed below.

In terms of wastewater characteristics, the BOD-to-TP ratio is high, with an average value of 55.1. This means that no additional food is required to support anaerobic phosphorus release. The BOD-to-TN ratio is high at 10, when 5 or greater would be recommended.

The plant uses a plug-flow, AS process with anoxic and aerobic basins in series. This was a retrofit design that the plant personnel implemented. Some unique features of this process are an anoxic basin with a long detention time, followed by a two-stage aerobic stage in series and, at the same time, the flexibility of operating parallel trains, such as during high-flow periods. The base mode of operation includes a long detention time at the anoxic basin (1 MG in basin 5), followed by an equal-size first aerobic basin (1 MG, basin 4 or 6) and then a smaller basin (either basin 3 or basins 1 and 2 combined). The internal recirculation from aerobic zone to the anoxic zone in the head area is up to four times the influent flow rate.

A unique operational strategy developed at the plant calls for a low return activated-sludge (RAS) flow rate and a deep sludge blanket in the clarifiers. The clarifiers are operated with 3 to 4 feet of blanket, while RAS is maintained at only 10 to 25 percent of the flow rate. In addition, the controlling parameter is mixed-liquor suspended solids (MLSS), ranging between 1,700 mg/L in summer and 2,400 mg/L in winter. There is no separate tank for volatile fatty acid generation. This practice has proven to provide full nitrification and a significant degree of denitrification in the retrofitted AS process. The average nitrate-nitrogen in the secondary effluent was 4 to 8 mg/L, leaving the denitrification filter to polish the effluent.

The plant uses denitrification filters manufactured by Leopold with a down-flow pattern and an automated system to control the methanol feed. The package includes a nitrate probe by Hach and a dosage-control algorithm by Leopold. The process is economical and efficient in denitrification. This is a compact process with a small footprint.

Another unique feature of this plant is that there is no primary settling and therefore all sludge produced is aerobic sludge. The sludge is pumped to the dewatering facility 5 miles away for dewatering with a cationic polymer. The filtrate is returned to the head of the plant for further processing.

Recycle loads are minimal because only aerobic digestion occurs on-site.

Wet-weather flows are managed with a normal mode of operation. The plant operated normally during a tropical storm in June 2006, when the flow increased from less than 4 MGD to more than 10.5 MGD in 3 days. Under extreme conditions such as a hurricane, the plant would shut down part of the aeration basin and protect the sludge inventory.

Cost Factors

Capital Costs

The main upgrades of the plant for biological nutrient removal (BNR) were implemented in 2000, when the existing aeration basins were reconfigured to allow an anoxic/anaerobic/aerobic series, and in 2005, when denitrifying filters were installed. The total cost for those upgrades, which were largely done by plant personnel, was $3.76 million. The components were updated to a total of $4.056 million in 2007 dollars using the *Engineering News-Record* Capital Cost Index (ENR CCI) index (USDA 2007).

It was assumed that 50 percent of the 2000 upgrade and 12 percent of the 2005 upgrade could be attributed to phosphorus removal, while 50 percent of the 2000 upgrade and 88 percent of the 2005 upgrade could be attributed to nitrogen removal. This attribution of the 2005 upgrade was based on the bulk of those capital improvements being for the denitrifying filter. The capital expenditure in 2007 dollars that could be attributed to phosphorus removal was $889,000. The annualized capital charge (20 years at 6 percent) was $77,500 for phosphorus removal.

The capital expenditure in 2007 dollars that could be attributed to nitrogen removal was $2.4 million. The annualized capital charge (20 years at 6 percent) was $210,000 for nitrogen removal.

The total capital attributed to BNR in 2007 dollars was $4.056 million. For the 7-MGD facility, the capital expenditure for BNR was $0.58/gpd capacity.

Operation and Maintenance Costs

The plant uses biological phosphorus removal to achieve the limit, while using methanol addition to complete the nitrogen removal. This means that the costs for phosphorus removal are all electrical, while the costs for nitrogen removal are electrical plus methanol. A summary of the electrical calculations is provided in an attachment at the end of this case study. The total electrical usage for phosphorus removal, assumed to be 30 percent of the total used, was 1,842,000 kilowatt-hours per year (kWh/yr). When the average electrical rate of $0.056/kWh was applied, the cost for phosphorus removal was $103,000 for the year. The total electrical usage for nitrogen removal was 4,170,000 kWh/yr, or $233,000.

The plant adds methanol at the rate of 83.1 gpd, at a cost of $1.75/gallon. This is equivalent to $53,000/yr for nitrogen removal.

Because of the methanol addition, an incremental amount of sludge is generated. The volume of methanol added is equivalent to 547 lb/day after accounting for the density of methanol, which is 0.79 g/cm^3. The chemical oxygen demand (COD) of the methanol is 1.5 lb COD/lb methanol, and the yield of volatile suspended solids (VSS) on methanol was assumed to be 0.4 lb VSS/lb COD (McCarty et al. 1969). The plant generated 328 lb sludge/day from methanol addition, or 59.9 ton sludge/yr. Assuming $200/ton for sludge disposal, the incremental amount for sludge addition attributed to nitrogen removal is $12,000.

Unit Costs for Nitrogen and Phosphorus Removal

During the evaluation period, the plant removed 69,900 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal is $1.48, while the unit capital cost is $0.73/lb of phosphorus removed.

During the same period, the plant removed 619,000 lb of TN. With the results above, the unit O&M cost for TN removal is $0.49, while the capital cost is $0.49/lb of TN removed.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal is $2.21/lb phosphorus removed, the life-cycle cost for ammonia nitrogen removal is $1.02/lb nitrogen removed, and the life-cycle cost for TN removal is $0.98/lb TN removed.

Cost-Effectiveness of the Denitrification Filter

The cost-effectiveness of the denitrification filter was evaluated separately for this plant. From filter influent and effluent data collected during a filter stress test in 2007, the filter on the average removes 3.5 mg/L nitrate-nitrogen. At a flow rate of 4.12 MGD, the filter removed 43,900 lb of nitrate-nitrogen during a year. Using the costs established above—$53,000 for methanol for the year and $12,000 for additional sludge disposal costs from methanol addition—the O&M cost per pound of nitrate removed in the denitrification filters is $65,000/43,900 = $1.48/lb nitrate-nitrogen removed.

Assessment of Magnitude of Costs and Main Factors

The life-cycle costs for phosphorus removal and full nitrification are extremely low, considering the phosphorus reduction level the plant has achieved. The main factors contributing to this achievement are the maximum use of existing facilities, good biological phosphorus removal, and efficient control with automation and many online sensors.

Assessment of magnitude of costs and main cost factors: The magnitude of cost at this facility is very low, mainly because of the availability of existing facilities and the original operating strategies of the plant personnel in maximizing both nitrogen and phosphorus removal at the retrofitted AS process. The new denitrification filters, therefore, use a minimal amount of methanol. In addition, no chemical is used to remove phosphorus. These factors make both the capital cost and O&M costs of this plant very low.

Discussion

Reliability factors: The plant achieves excellent performance at the mean concentration of 2.14 mg/L of TN with a COV of 16 percent. This is mainly because the plant has two separate-stage denitrification processes with an external carbon source at the second stage, or dentrification filter. Operational strategies developed by the plant personnel achieved a significant amount of denitrification in the AS process, followed by a separate-stage polishing with an automated feed strategy using an online nitrate probe. For phosphorus removal, the mean concentration of 0.26 mg/L is excellent, while the COV is moderate at 62 percent. This low a level is remarkable for an entirely biological phosphorus removal process. Note that the denitrification filter by Leopold uses a down-flow process and therefore removes suspended solids concurrently with nitrogen removal.

Cost factors: Three key factors are identified in achieving a high level of BNR at a low cost at this facility: (1) the maximum use of an existing AS process with minimal retrofit costs; (2) development of an original operating strategy to maximize BNR in the retrofitted AS process; and (3) a separate-stage denitrification with minimal methanol feeding. This combination of biological phosphorus removal and a down-flow denitrification filter in series resulted in a reliable, low-cost solution for both nitrogen and phosphorus removal.

Summary

This facility removes both nitrogen and phosphorus exceptionally well and reliably. The two-stage biological processes in series offer the highest efficiency in nutrient removal at minimum costs. The source of wastewater is typical residential customers in the suburb of a large metropolitan area. The BOD-to-TP ratio averages 55.1. The retrofitted AS process consists of an anoxic stage with a 4.8-hour residence time, followed by an aerobic stage in two tanks with a residence time of 11.5 hours. The operating strategy developed at this facility is unique because the sludge blanket at the clarifiers is 3 to 4 feet deep and the RAS flow rate is maintained at a low (10–25 percent) portion of the plant flow. The second-stage denitrification filters then remove the remaining nitrogen with a methanol feed.

The design and operation result in a high level of removal—an effluent TN concentration of 2.14 mg/L with a COV of only 19 percent and an effluent TP concentration of 0.26 mg/L with a COV of 62 percent.

The costs of removal were very low for both capital and O&M. The life-cycle cost for removal of TP was $2.21/lb of TP removed, while the life-cycle cost for TN removal was $0.98/lb of TN removed, including the cost for methanol. The capital cost for the flow capacity was low at $0.58/gpd capacity.

Acknowledgments

The authors are grateful for the significant assistance and guidance provided by Haywood Phthisic, III, director of Utilities in Johnston County, North Carolina. This case study would not have been possible without Mr. Phthisic's prompt response, with well-deserved pride in the facility and its operation. Thanks are extended to Johnston County for participating in this case study for the U.S. Environmental Protection Agency.

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Attachment: Electrical Cost Calculation

Electrical									
Anoxic/Anaerobic Mixers									
HP	Number	Power draw (kW)	kWh/day	kWh draw/day	kWh	%P	%N	For P	For N
10	3	22.38	24	537.12	196,048.8	70	30	137234.2	58,814.64
15	1	11.19	24	268.56	98,024.4	70	30	68617.08	29,407.32
15	1	11.19	24	268.56	98,024.4	70	30	68617.08	29,407.32
Blowers									
150	2	223.8	24	5,371.2	1,960,488	30	70	588146.4	1,372,342
100	2	149.2	24	3,580.8	1,306,992	30	70	392097.6	914,894.4
Filter Pumps									
150	3	335.7	24	8,056.8	2,940,732	20	60	588146.4	1,764,439
Total Draw					6,600,310			1,842,859	4,169,304
Methanol	83.1	gal/day							
	1.75	cost/gal							
	145.425	cost/day							
	53,080.125	cost/yr							

Fiesta Village Advanced Wastewater Treatment Plant
Lee County, Florida
Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

This plant was selected as a case study because it is a good example of the use of the denitrification filter process. The plant consists of an extended air oxidation ditch process followed by denitrification filters with methanol feed. Phosphorus removal is achieved with alum feed to the secondary effluent. Nitrogen and phosphorus are being removed successfully down to 3 and 0.1 milligrams per liter (mg/L), respectively.

The Fiesta Village Advanced Wastewater Treatment Plant is in Lee County, Florida. It is permitted for 5 million gallons per day (MGD) capacity, and in 2006 it processed an average of 3.16 MGD. The plant is designed to send 2.0 MGD (annual average) into a slow-rate, public-access reuse system for irrigation of golf courses and residential developments. It has the potential for future reuse expansion to 3.158 MGD. Any water not reused, including stormwater flow, is permitted for a surface water discharge to the Caloosahatchee River.

The relevant National Pollutant Discharge Elimination System (NPDES) permit limits for the facility are shown in Tables 1 and 2.

Table 1. NPDES permit limits

Parameter (mg/L unless stated)	Annual average	Monthly average	Weekly average	Daily maximum
BOD_5	20	25	40	60
TSS	20	30	45	60
Total nitrogen	3	3	4.5	6
Total phosphorus	0.5	0.5	0.75	1

Notes:
BOD = biochemical oxygen demand.; TSS = total suspended solids

Table 2. Reuse water permit limits

Parameter (mg/L unless stated)	Annual average	Monthly average	Weekly average	Daily maximum
BOD_5	20	30	45	60
TSS				5
Residual chlorine				1 (minimum)

Treatment Processes

Figures 1 and 2 present the plant layout and process flow for the Fiesta Village Facility. The plant is an extended-aeration oxidation ditch facility, and the treatment process includes an odor control system, primary bar manual/mechanical screening, aerated grit removal, two oxidation ditches, two clarifiers, two aerobic digesters, three screw lift pumps, four denitrification filters, dual chlorine contact chambers, effluent transfer pumping station, chemical feed equipment, sulfur dioxide dechlorination, post-re-aeration, a reuse storage tank, and a high-service reuse/effluent pump station.

Basis of Design and Actual Flow

The design flow for the facility is 5 MGD. The average flow for the study period was 3.16 MGD, while the maximum month flow during the study period was 4.14 MGD during July 2006. The peak day flow recorded was 5.78 MGD.

Design loadings:
- Biochemical oxygen demand (BOD): 240 mg/L
- Total suspended solids (TSS): 268 mg/L
- Total Kjeldahl nitrogen (TKN): 37 mg/L
- Total nitrogen (TN): 38.2 mg/L
- Total phosphorus (TP): 7.3 mg/L
- Alkalinity: 284 mg/L as calcium carbonate ($CaCO_{3)}$

Oxidation ditch—437 ft long x 80 ft wide x 12 ft deep, or 3 million gallons (MG), each
- Anoxic zone: one aerator turned off, or 25 percent by volume
- Aerators: 60 hp, four each per oxidation ditch
- Hydraulic retention time (HRT): 28.8 hours
- Mixed liquor suspended solids (MLSS): 3,500 mg/L
- Mixed liquor volatile suspended solids (MLVSS): 2,500 mg/L
- Mean cell residence time: 30 days
- Food to microorganism ratio: 0.1:0.4 lb BOD/lb MLVSS
- Waste activated sludge (WAS): 0.06 MGD, each, or 6,500 lb/day, each
- Dissolved oxygen (DO): 0.5–2.0 mg/L in aerobic zone and 0.1–0.5 mg/L in anoxic zone

Secondary clarifiers—diameter of 90 ft (each, and there are two)
- Volume: 0.665 MG (each)
- Surface area: 5,538 ft^2 (each) and surface loading rate = 600–1,200 gpd/ft^2
- Blanket depth: less than 3 ft

Return activated sludge (RAS)—rate at 100 percent of plant influent, or 3.5 MGD (3 each)

Denitrification filter—10 ft x 40 ft, 4 cells each
- Hydraulic loading rate: 2.2 gpm/ft^2 at design

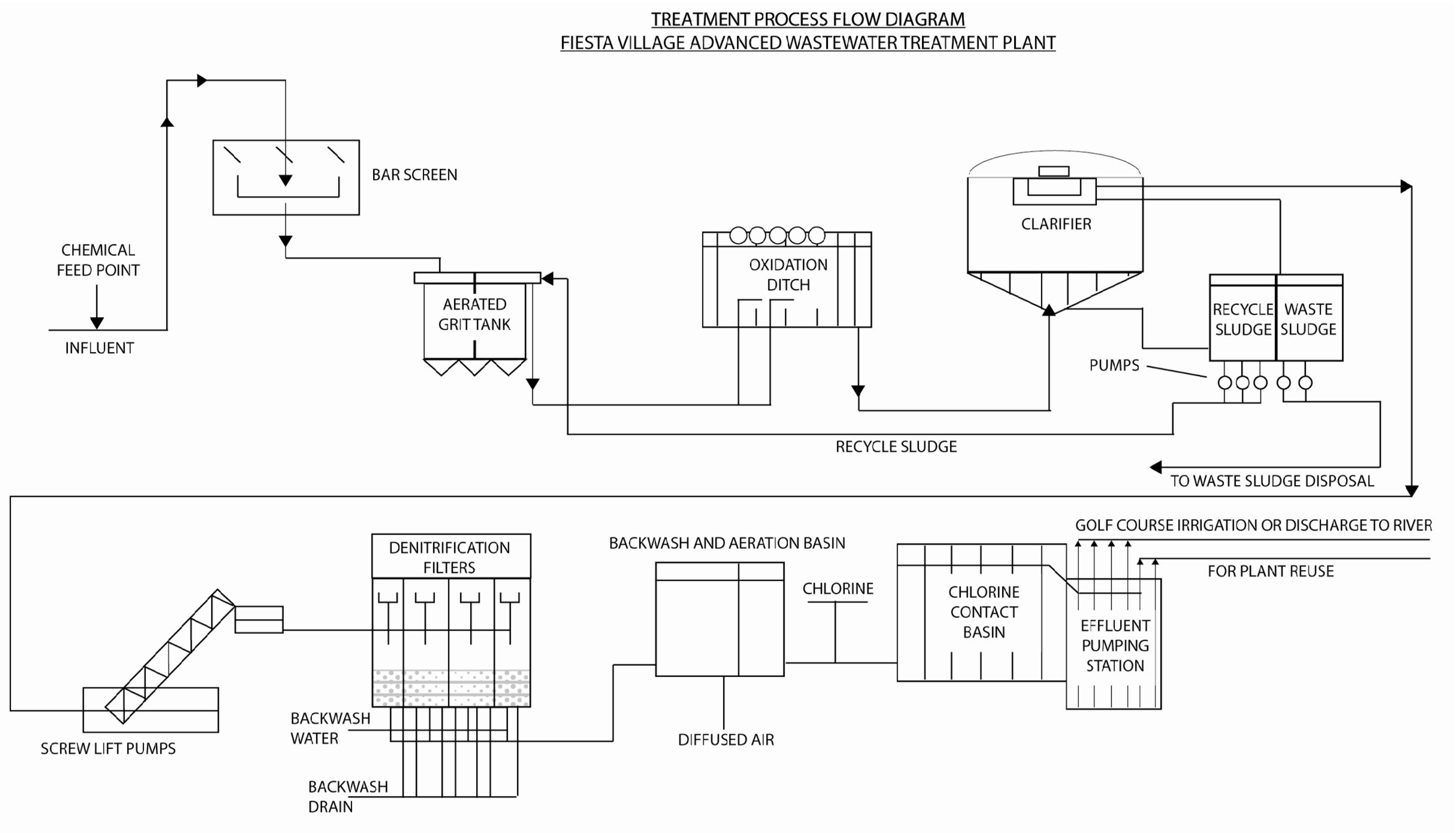

Figure 1. Treatment process flow diagram Fiesta Village Advanced Wastewater Treatment Plant.

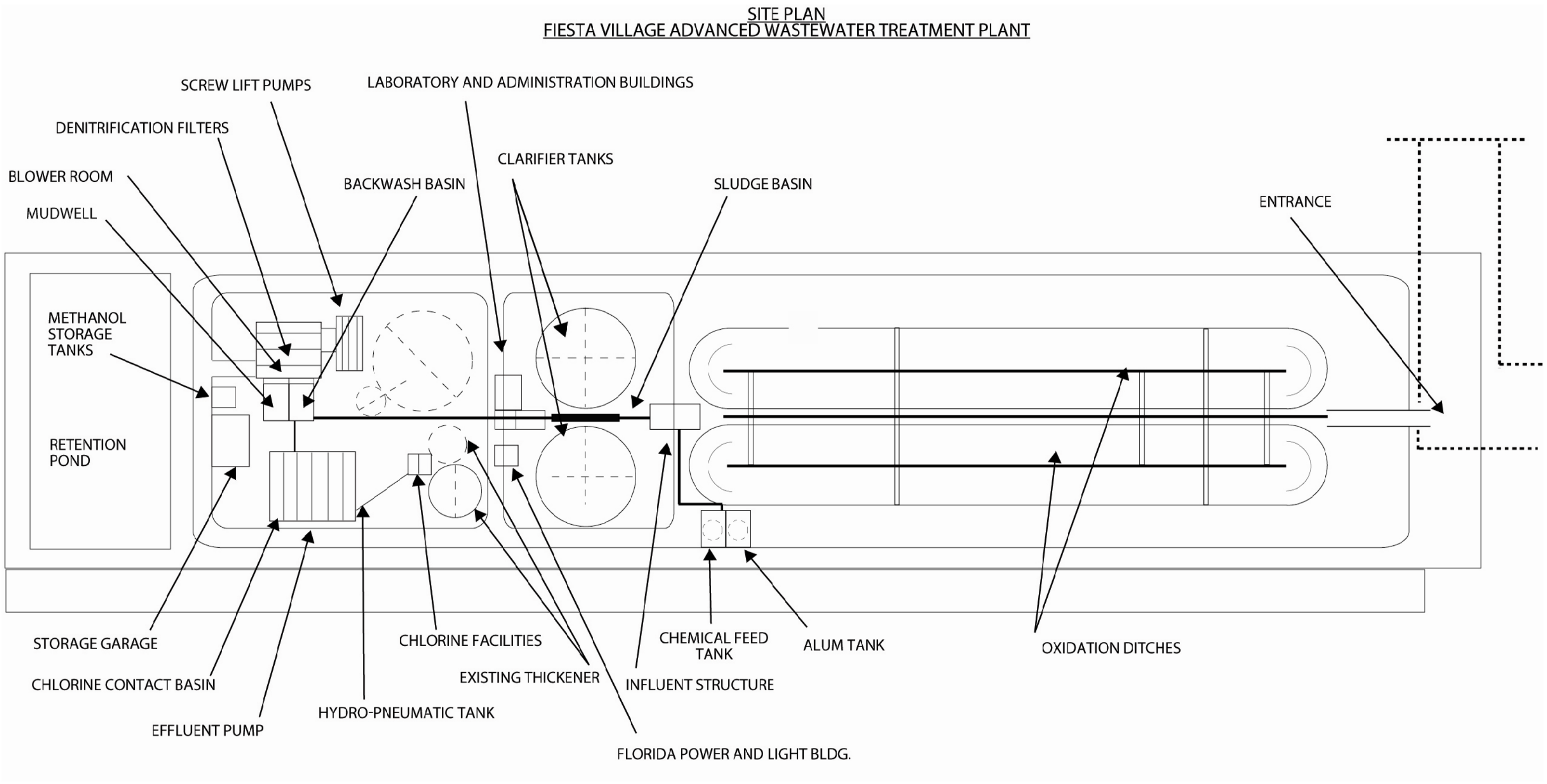

Figure 2. Site plan Fiesta Village Advanced Wastewater Treatment Plant.

Aerobic Digestion

- Diameter: 39 ft, 16 ft deep
- Volume: 0.143 MG, 2 each
- Disc diffusers
- Loading rate: 0.01–0.02 lb VSS/ ft^3 day
- DO: 1–3 mg/L
- Sludge age: 5–40 days
- Digester temperature: less than 30 degrees Celsius (°C)

Plant Parameters

Overall plant influent and effluent average results for the period January 2006 to December 2006 are shown in Table 3.

Table 3. Fiesta Village influent and effluent averages

Parameter (mg/L unless stated)	Average value	Maximum month	Max month vs. ave.	Maximum week	Sample method/frequency
Flow (MGD)	3.16	4.14	31%	4.26	
Influent TP	3.85	4.58	18%	--	Monthly/composite
Effluent TP	0.102	0.19	85%	0.39	Daily/composite
Influent BOD	134	167	24%	179	Daily/composite
Effluent BOD	1.37	2.95	116%	5.2	Daily/composite
Influent TSS	199	261	31%	348	Daily/composite
Effluent TSS	0.72	1.17	61%	1.48	Daily/composite
Influent NH_4-N	27.2	34.5	27%	--	Monthly/composite
Effluent NH_4-N	0.13	0.2	50%	0.28	Daily/composite
Secondary Effluent NO_3-N	2.9[a]	3.0[a]	7%	3.9[a]	Daily/composite
Influent TN	33.2	50.6	53%	--	Monthly/composite
Effluent TN	1.71	2.61	53%	3.90	Daily/composite

Notes:
BOD = biochemical oxygen demand
Max month vs. average = (max month – average)/average x 100
NH_4-N = ammonia measured as nitrogen
TN = total nitrogen
TP = total phosphorus
TSS = total suspended solids
[a] Jan–April 2007

Table 4 presents plant monthly averages for the process parameters, as available.

Table 4. Monthly averages for plant process parameters

Month	MLSS (mg/L)	Sludge age (d)	HRT (hr)	Temperature (°C)
Jan 2006	3,578	37	48	--
Feb 2006	3,807	39	44	--
Mar 2006	4,085	35	46	--
Apr 2006	3,845	24	50	--
May 2006	3,510	33	55	--
June 2006	3,564	28	47	--
July 2006	3,571	32	35	30.4
Aug 2006	3,480	36	44	--
Sept 2006	3,495	34	39	--
Oct 2006	3,509	37	49	--
Nov 2006	3,775	59	49	--
Dec 2006	4,204	41	50	--

Performance Data

Figure 4 presents reliability data for the removal of TP. The removal is good, with an effluent TP average of 0.1 mg/L and a medium coefficient of variation (COV) of 35 percent.

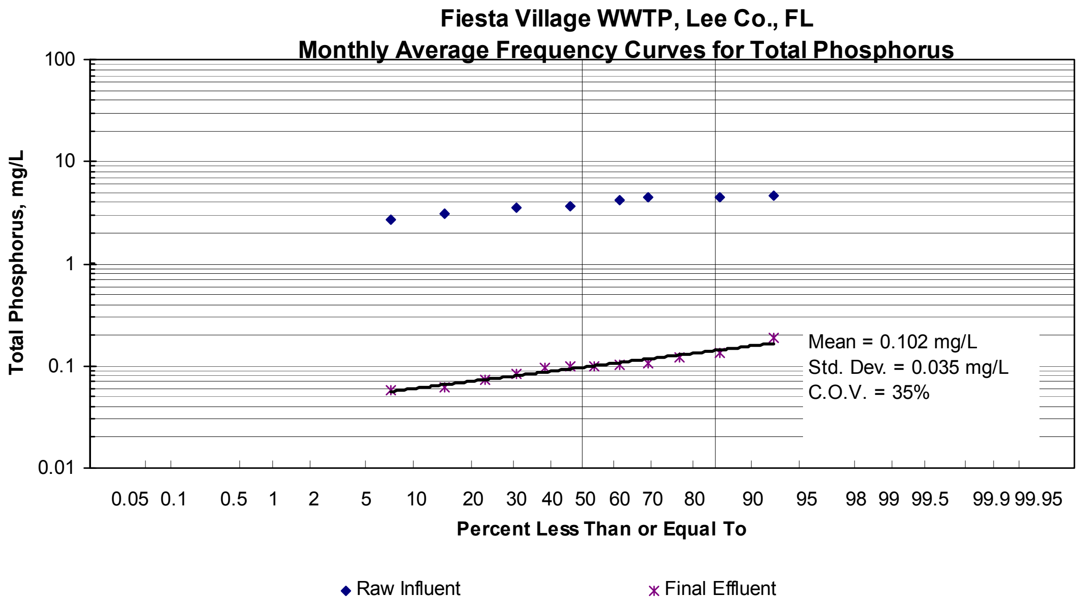

Figure 4. Monthly average frequency curves for TP.

Figure 5 presents reliability data for ammonia nitrogen removal. The removal of ammonia nitrogen is very good, with a mean effluent of 0.134 mg/L and a low COV of 40 percent.

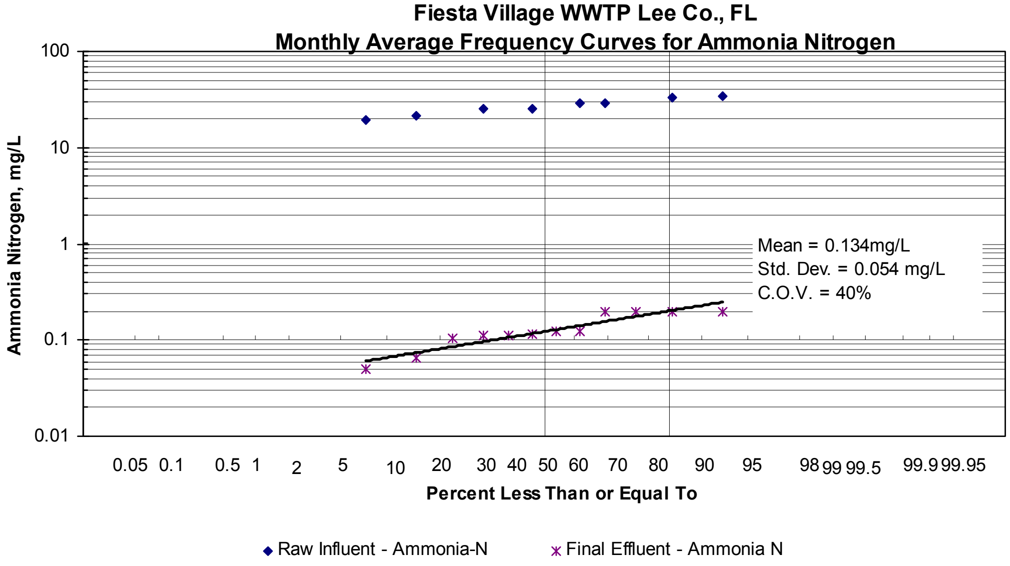

Figure 5. Monthly average frequency curves for ammonia nitrogen.

Figure 6 present reliability data for removal of TN. Nitrogen is removed in two steps at this facility. The oxidation ditch takes nitrate-nitrogen down to an average of 3 mg/L, and then the denitrification filter takes it down to an annual average of 1.45 mg/L. at a COV of 28 percent.

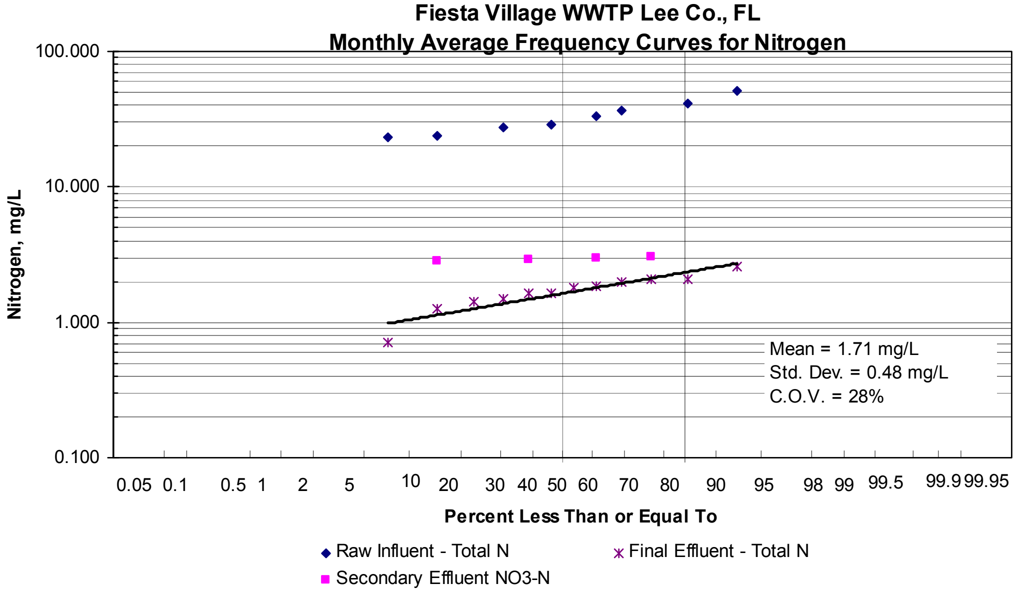

Figure 6. Monthly average frequency curves for nitrogen.

Reliability Factors

This facility is unique in three ways: separate-stage denitrification using methanol; alum feed to the oxidation ditch effluent prior to the secondary clarifiers for chemical phosphorus removal; and filtration of effluent with the same denitrification filters. The facility is also unusual in that it has no primary settling and thus all sludge generated is kept aerobic before it is disposed of off-site at another county facility.

The results are excellent. The plant achieved a TN concentration of 1.71 mg/L with a COV of 28 percent and a total phosphorus (TP) concentration of 0.1 mg/L with a COV of 35 percent. The key factors contributing to this performance are described below.

The key reason for excellent denitrification is the use of two processes in series—the first in the oxidation ditch for most of the removal, followed by polishing at the denitrification filter. The oxidation ditch is operated with the target nitrate-nitrogen concentration of 3.0 to 3.5 mg/L and ammonia nitrogen at 0.2 mg/L in the secondary effluent. This target removal is accomplished under the current loading conditions by turning one of four brush aerators off during the day and two off during the night, thereby maintaining 25 percent and then 50 percent of the volume, respectively, as an anoxic zone. The DO concentration in the oxidation ditch is adjusted using the remaining brush aerators. The oxidation ditch is operated with a long SRT (30–40 days) and HRT (20–30 hours). In addition, another unique operating plan includes the denitrification blanket in the clarifiers. The sludge blanket depth is maintained at between 2.5 and 3.5 feet.

The denitrification filters then brings the nitrate-nitrogen to below 2 mg/L, with a low methanol feed rate of 129 lb per day. The methanol–to–nitrate-nitrogen ratio averaged 1.9 pounds of methanol per pound of nitrate present, or 2.4 lb per pound of nitrate removed. The plant measures nitrate-nitrogen in the effluent in adjusting the methanol feed rate, which is steady year-round.

Alum was fed at the average dosage of 8.9 mg/L as aluminum, or at the aluminum-to-TP ratio of 2.31, in achieving a low concentration of 0.1 mg/L for the year.

Recycle loads are minimal at this facility because aerobically digested sludge is hauled away to another facility for final sludge processing.

During wet-weather periods, a normal mode of operation is maintained. Under extreme peak flow conditions, the clarifiers are protected from surges by shutting off a number of brush aerators.

Costs

Capital Costs

The main upgrades of the plant for biological nitrogen removal (BNR) occurred in 1984, when Phase 1, consisting of the east oxidation ditch, east clarifier, denitrifying filter, and major structures for the west ditch and west clarifier were installed; in 1986, when Phase 2 improvements were installed; and in 2002, when equipment for the west oxidation ditch and west clarifier was installed. Table 5 presents the costs for those improvements (Voorhees et al. 1987; TKW Online 2007), along with capital cost updates based on the *Engineering News-Record* Capital Cost Index (ENR CCI). The ENR CCI, compiled by McGraw-Hill, provides a means of updating historical costs to account for inflation, thereby allowing comparison of costs on an equal basis. From a Web site provided by the U.S. Department of Agriculture (USDA 2007), the ENR index for 1984 was 4,146; for 1986, 4,295; for 2002, 6,538; and for May 2007, 7,942.

Table 5. Plant improvement costs

	Year	Original cost	2007 cost	%P	%N	%other	P cost	N cost
Phase 1	1984	$6,505,833	$12,462,452	2%	50%	48%	$249,249	$6,231,226
Denite Filter	1984	$930,059	$1,781,604	12%	88%	0%	$213,792	$1,567,811
Controls	1984	$441,323	$845,390	2%	50%	48%	$16,908	$422,695
Phase 2	1986	$1,200,000	$2,218,952	0%	50%	50%	$0	$1,109,476
Phase 3	2002	$6,800,000	$8,260,263	0%	50%	50%	$0	$4,130,132
TOTAL			$25,568,661	--	--	--	$479,949	$13,461,340

The table also shows the percentage of capital cost for each unit that was attributed to phosphorus or nitrogen removal; the rest of the capital cost was attributed to other treatment, particularly biochemical oxygen demand (BOD) and total suspended solids (TSS) removal and disinfection. Because the plant is not doing biological phosphorus removal, it was assumed that only 2 percent of the Phase 1 cost plus 2 percent of the cost of controls could be attributed to phosphorus removal for the alum addition system. Because the denitrification filters remove solids, including aluminum phosphate precipitate, it was assumed that 12 percent of that cost could be attributed to phosphorus.

On the basis of DO usage, it was assumed that 50 percent of the cost of Phases 1, 2, and 3 could be attributed to nitrogen removal. It was assumed that 88 percent of the cost of the denitrification filters could be attributed to nitrogen removal. To be consistent with other case studies in this document, it was assumed that 50 percent of the control costs could be attributed to nitrogen removal.

The above analysis resulted in a total of $480,000 in capital attributed to phosphorus removal and $13,461,000 attributed to nitrogen removal, in 2007 dollars. The annualized capital charge for phosphorus removal (20 years at 6 percent) was $42,000. The annualized capital charge for nitrogen removal was $1,174,000.

The total capital attributed to nutrient removal, in 2007 dollars, was $13.9 million. For the 5-MGD facility, this means the capital expenditure per gallon of treatment capacity was $2.79.

Operation and Maintenance Costs

The plant uses chemical phosphorus removal and BNR, with extensive use of alum for the former and methanol as a supplemental carbon source for the latter. This means that the cost for phosphorus removal is essentially all for chemicals and for the disposal of the resulting sludge, while the cost for nitrogen removal is electrical (for the aeration basins), chemical (for the methanol), and for the disposal of the extra sludge resulting from methanol addition. A summary of the electrical calculations is provided in the Attachment. It was assumed that some of the electricity for the blowers could be attributed to phosphorus removal, to account for mixing alum in the ditch. The total electrical usage for nitrogen removal was 1,911,000 kilowatt-hours (kWh). When the average electrical rate of $0.12/kWh (including demand charges) was applied, the cost of electricity for nitrogen removal was $229,000.

Alum is applied for both phosphorus removal and TSS reduction to meet the permit requirements for water reuse. The average amount of alum applied over the period was 151 gallons/MG of flow; assuming $0.66/gallon, the cost of alum was $115,400. It was assumed that 30 percent of the alum cost was attributed to phosphorus removal, bringing the chemical cost for phosphorus removal to $34,600.

Methanol is applied at the denitrification filter to promote nitrate removal. The total amount of methanol added over the study period was 47,000 lb. Assuming a cost of $0.27/lb (cost of methanol for another case study plant), the chemical cost for nitrogen removal was $12,500.

The alum added (8.9 mg/L as Al) was assumed to entirely convert to aluminum hydroxide sludge; at the average flow of 3.16 MGD, this was 677 lb of aluminum sludge per day, or 124 dry tons/year. The plant trucks its sludge at an average cost of $0.048/gallon. Assuming a concentration of 2 percent solids, the 124 dry tons of alum sludge is equivalent to 1,486,000 gallons of sludge. Assuming 30 percent of the sludge is associated with phosphorus removal, the cost for phosphorus sludge disposal was $21,700.

The 47,000 lb/yr of methanol has a chemical oxygen demand (COD) of 1.5 lb COD/lb methanol, or 70,750 lb COD/yr. The typical yield of volatile suspended solids (VSS) on methanol is 0.4 lb VSS/lb COD, giving 28,300 lb sludge/yr, or 14.2 tons sludge/yr from methanol addition. At a solids concentration of 2 percent, this means an additional 708 tons, or 170,000 gal/yr, of liquid sludge to haul to other Lee County plants for treatment and disposal. The total hauled during 2006 was 7,520,000 gallons, meaning the methanol sludge was approximately 2.2 percent of the total. At the plant's average disposal charge of 4.9 cents/gallon, the total cost for nitrogen removal sludge was $8,300.

Unit Costs for Nitrogen and Phosphorus Removal

During the evaluation period, the plant removed 36,100 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal is $1.77, while the unit capital cost is $1.16/lb of phosphorus removed. If the plant were operating at full capacity (5 MGD), the unit O&M cost for phosphorus removal would be $1.34, with the unit capital cost $0.73/lb of phosphorus removed.

During the evaluation period, the plant removed 303,000 lb of TN. With the results above, the unit O&M cost for nitrogen removal is $0.91, while the capital cost is $3.87/lb of TN removed. If the plant were operating at full capacity, the unit O&M and capital costs would be $0.57 and $2.45, respectively, per pound of TN removed.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal is $2.93/lb of phosphorus removed, the life-cycle cost for TN removal is $4.78/lb of TN removed, and the life-cycle cost for ammonia nitrogen removal is $5.57/lb of nitrogen removed. For full-capacity operations, the costs would be $2.07/lb for phosphorus, $3.02/lb for TN, and $3.52/lb for ammonia nitrogen.

Assessment of magnitude of costs: The capital cost of $2.79 per gpd capacity is on the high side, but the O&M costs are moderate because of the low electrical costs but high chemical costs.

Discussion

Reliability factors: The performance has been very reliable in nitrogen and phosphorus removal. Nitrogen removal was achieved very reliably by having two processes in series for denitrification. Most of the removal was accomplished by the optimal use of the oxidation ditch system, where denitrification was achieved in anoxic zones of various sizes, as well as in the denitrifying sludge blanket in the clarifiers. The polishing of nitrate was accomplished at the denitrification filters with minimal dosage of methanol. Phosphorus removal was accomplished by alum addition before the secondary clarifiers, followed by the same denitrification filters, making the process both efficient and reliable.

Cost factors: Costs for both methanol and alum are low because of the optimal use of the existing facilities. Costs are also low because sludge is not processed on-site.

Summary

The Fiesta Village facility is an advanced wastewater treatment plant with an oxidation ditch followed by secondary clarifiers and four dedicated denitrification filters. The performance was highly efficient and reliable for the year studied. Nitrogen removal was achieved biologically to the mean concentration of 1.44 mg/L with a COV of 27 percent. Many factors contributed to this high result, including maximum use of the oxidation ditch for denitrification, thereby reducing the load to the denitrifcation filters. The personnel at the facility are credited for developing daily operating procedures for the control parameters and implementing them consistently. Using denitrifying blankets in the clarifiers and maintaining flexible anoxic zones in the oxidation ditch are two unique features of the operation in achieving effluent nitrate-nitrogen concentration of 3 mg/L as a monthly average. The methanol usage was minimal at the average dosage of 1.9 lb per pound of nitrate applied, compared to 3 lb in the literature.

Acknowledgments

The authors of this report acknowledge with gratitude the significant assistance and guidance provided by Tom Hill, Lee County utilities deputy director; Dennis Lang, chief operator at the Fiesta Village Facility; and Jon Meyer, Utilities Operations Manager. This report would not have been possible without their prompt response with well-deserved pride in their facility and operation. EPA acknowledges Lee County, Florida, for its participation in this case study.

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Attachment: Electrical and Chemical Costs

Electrical								for P	for N
Hp	Number	kW Power draw	hours/day	kWh draw/day	kWh draw/year	%P	%N		
Aerator 60	8	358.08	24	8593.92	3136781	2	50	62735.6	1568390.4
RAS pump 30	3	67.14	24	1611.36	588146.4	0	50	0	294073.2
WAS pump 7.5	2	11.19	24	268.56	98024.4	0	50	0	49012.2
Total draw					3822952			62735.6	1911475.8
Alum cost	$115,338								
% for P removal	30								
Alum cost for P	$34,616								
Methanol cost	$12,735	(all for N removal							

Kalispell Advanced Wastewater Treatment
Kalispell, Montana

Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

The Kalispell Wastewater Treatment Plant (WWTP) is an advanced wastewater treatment facility in Kalispell, Montana. Kalispell is in the northwestern part of the state, near Glacier National Park. The area is subjected to extreme weather conditions, with temperatures ranging from 95 degrees Fahrenheit (^{o}F) in the summer to –30 ^{o}F in the winter.

This facility was selected as a case study because of good biological phosphorus removal and nitrification using a modified University of Cape Town (UCT) process with the fermenter technology in a cold region.

The facility began operating in October 1992 to protect Flathead Lake, the largest freshwater lake west of the Mississippi River. The plant has received a national first place and two Region 8 first place Operations and Maintenance Excellence Awards from the U.S. Environmental Protection Agency (EPA), a Commendation of Excellence Award from the Flathead Basin Commission, and a System of the Year Award from Montana Rural Water Systems. In addition, the processes for nitrogen removal was designed and implemented as a voluntary initiative.

Kalispell has experienced a significant increase in population since the facility was constructed. The city plans to expand the plant over the next several years to accommodate growth. The expansion will add to or replace some units and modify others to continue the concept of treatment without using chemicals. The plant is designed with expansion planned for the flows and loads shown in Table 1.

Table 1. Design flow and loads

Year	Flow (MGD)	BOD_5 (mg/L)	TSS (mg/L)	TKN (mg/L)	TP (mg/L)
2000	2.5	216	259	25	4.5-6.5
2008	3.0	216	260	25	4.5-6.5

Notes:
BOD_5 = biochemical oxygen demand
MGD = million gallons per day
TKN = total Kjeldahl nitrogen
TSS = total suspended solids
TP = total phosphorus

The National Pollutant Discharge Elimination System (NPDES) permit limits for the plant are shown in Table 2.

Table 2. NPDES permit limits

Parameter	7-day average (mg/L)	30-day average (mg/L)
BOD_5	15	10
TSS	15	10
Total P	--	1.0
Ammonia nitrogen	--	1.4 (sufficient to meet stream limits)

Treatment Processes

Wastewater treatment at the Kalispell WWTP begins with flow entering the plant through a 36-inch-diameter pipe from the city’s system. The influent flows through the headworks and is pumped to two rectangular primary clarifiers by five low-head lift pumps. Primary clarifier effluent then flows into the bioreactor, which consists of 11 tanks in series. During periods of high flow, primary effluent is directed to the equalization basin. Flow from the equalization basin is then returned to the primary clarifiers during periods of lower influent flow.

The system at Kalispell is unique because it is based on the modified UCT process with additional flexibility provided by swing zones that can be operated in several different modes. Four zones (anaerobic, first and second anoxic, and aerobic) are created for solids and nutrient removal. Depending on the chemistry and biology, the plant personnel can determine the optimum number of anaerobic zones and, thus, the subsequent anoxic zones. Bioreactor effluent flows to two circular, center-drive secondary clarifiers and then through an effluent deep-bed sand filter, with an up-flow, continuous backwash design. The filtered effluent then flows through an ultraviolet disinfection system and is re-aerated before it is discharged to Ashley Creek.

The solids process train in the plant starts with the primary sludge that is removed from the primary clarifiers by two primary sludge pumps to the completely mixed fermenter. Primary sludge is pumped to the fermenter at timed intervals—typically at 4.8 minutes per hour. The target solids concentration in the fermenter is 12,000 milligrams per liter (mg/L). Waste fermented sludge flows to the gravity thickener; two pumps return the fermenter supernatant to the bioreactor. The fermenter has a volume of 118,000 gallons, a hydraulic retention time of 7 to 21 hours, and a mixing power of 0.06 horsepower (HP) per 1,000 gallons. The solids retention time (SRT) is designed to be 4 to 5 days.

Sludge from the gravity thickener is pumped to the primary digester and then to the two secondary digesters. Digested primary sludge is pumped to two belt filter presses. Secondary sludge is pumped as return activated sludge (RAS) to the bioreactor. The RAS is pumped by two RAS pumps to two dissolved air flotation (DAF) thickeners. DAF filtrate is wasted back to the bioreactor, and the thickened sludge from the DAF is pumped via two DAF float pumps to two belt filter presses, where it is mixed with digested primary sludge just before the presses. The DAF sludge is not anaerobically digested to avoid re-release of accumulated phosphorus. The belt press cake is trucked to a composting operation. Digester supernatant and the filtrate from belt press are returned to the headworks.

Figure 1 shows the overall process flow diagram. Figure 2 shows details of the biological reactor and how RAS can be directed to one of three cells depending on operating conditions. The fermenter supernatant also can be directed to any of the first three cells as conditions warrant.

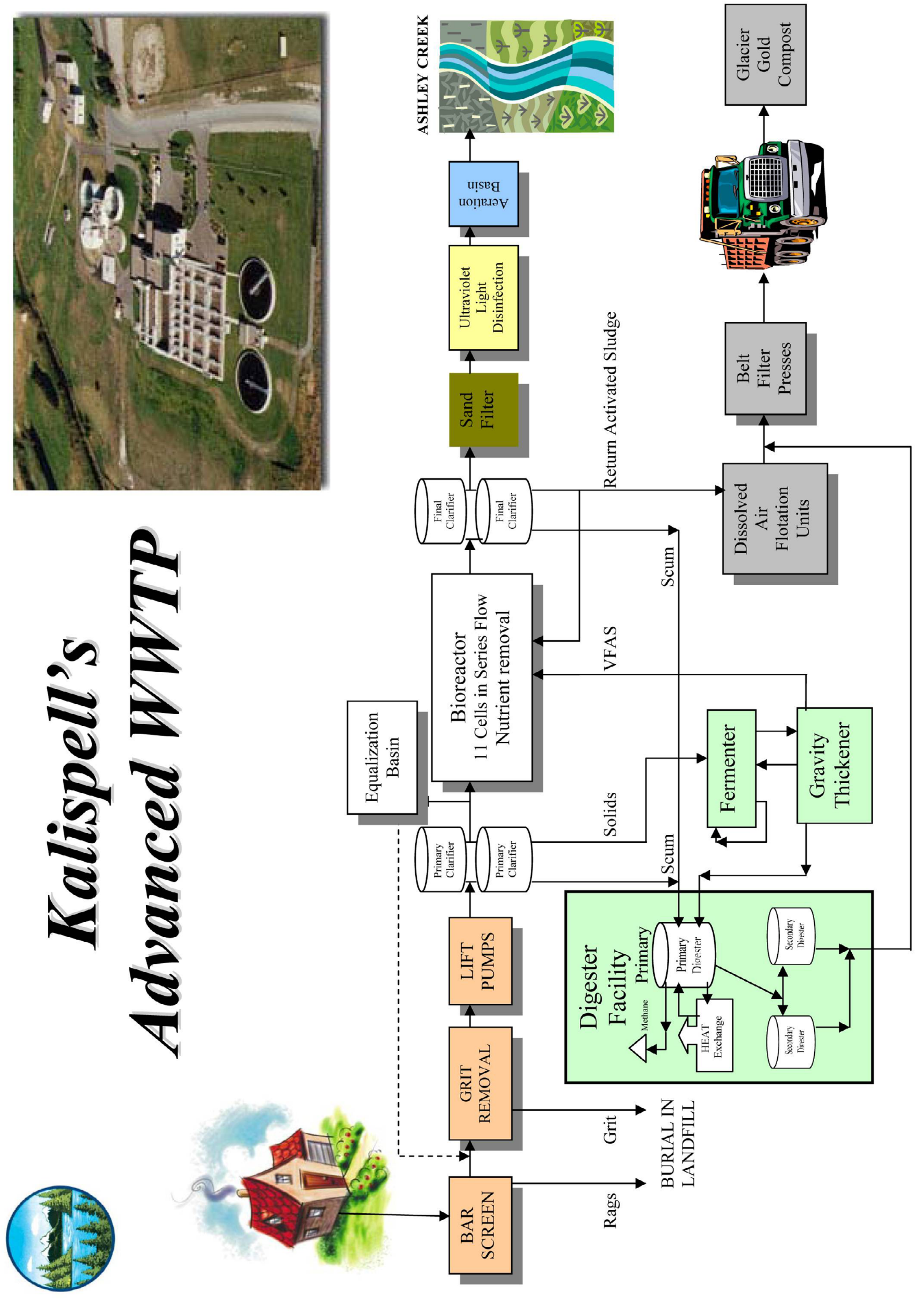

Figure 1. Kalispell's advanced WWTP.

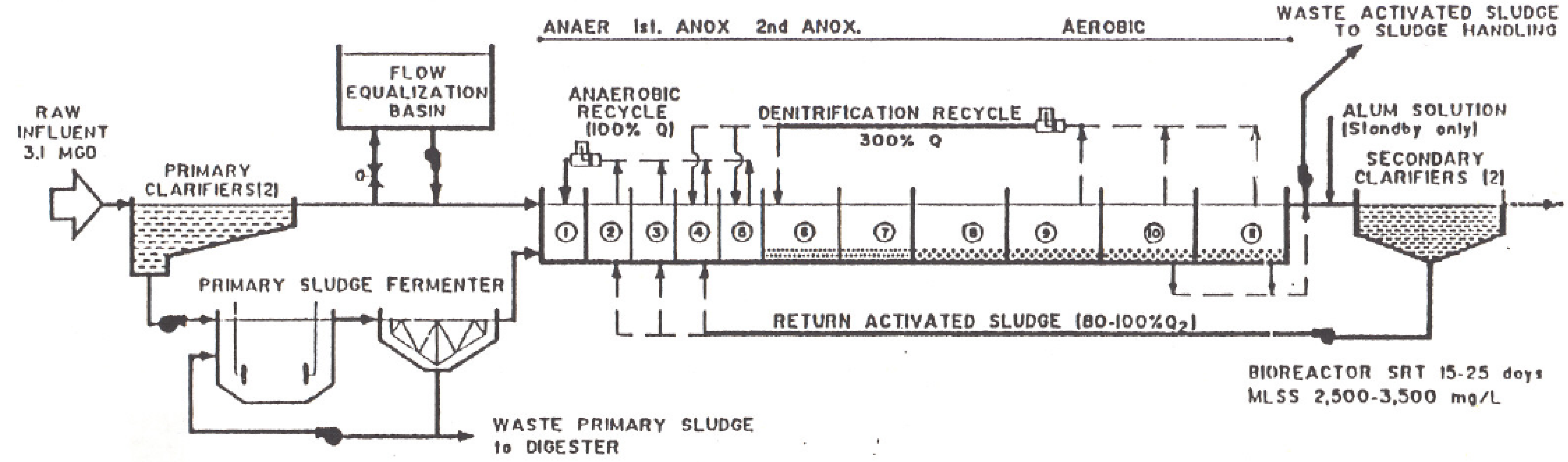

Figure 2. Kalispell biological reactor detail.

Plant Parameters

Overall plant influent and effluent average results for the period July 2005 through June 2006 are shown in Table 3.

Table 3. Influent and effluent averages

Parameter (mg/L unless stated)	Average	Maximum month	Max month vs. avg.	Maximum week	Sample method/frequency
Flow (MGD)	2.95	3.45	17%	4.04	--
Influent TP	4.11	4.88	19%	5.2	Composite/weekly
Effluent TP	0.12	0.15	25%	0.31	Composite/weekly
Influent BOD	226.36	282	25%	428	Composite/weekly
Effluent BOD	< 4	< 4	0%	5.8	Composite/weekly
Influent TSS	225.17	326	45%	680	Composite/weekly
Effluent TSS	1.21	2.9	140%	4.1	Composite/weekly
Influent NH_4-N	24.35	29.4	21%	--	Grab/monthly
Effluent NH_4-N	< 0.07	< 0.07	0%	--	Grab/monthly
Influent TKN	39.28	47	20%	--	Grab/monthly
Effluent TKN	0.63	1.26	100%	--	Grab/monthly
Influent TN	39.6	48.0	21%	--	Grab/monthly
Effluent TN	10.6	19.9	86%	--	Grab/monthly

Notes:
BOD = biochemical oxygen demand
Max month vs. average = (max month – average)/average x 100
MGD = million gallons per day
NH_4-N = ammonia measured as nitrogen
TKN = total Kjeldahl nitrogen
TN = total nitrogen
TP = total phosphorus
TSS = total suspended solids

Table 4 presents plant monthly averages for process parameters.

Table 4. Monthly averages for plant process parameters

Month	MLSS (mg/L)	Sludge age (d)	HRT (hrs)	Water temp (°C)
July 2005	2,586	10	13	18.9
Aug. 2005	2,517	8	14	20
Sept 2005	2,625	11	13	18.6
Oct 2005	2,659	12	14	17.1
Nov 2005	2,637	10	15	14.9
Dec 2005	2,808	11	15	12.1
Jan 2006	2,744	10	12	11.4
Feb 2006	2,757	9	13	10.8
Mar 2006	2,657	9	14	10.9
Apr 2006	2,568	9	11	12.3
May 2006	2,536	9	13	14.8
June 2006	2,529	9	11	16.7

Notes:
HRT = hydraulic retention time
MLSS = mixed liquor suspended solids

Performance Data

This section provides information about the operational performance of nutrient removal at the plant. Figures 3 and 4 present reliability plots for monthly average and weekly average phosphorus. For the monthly average data, the facility has a very low coefficient of variation (COV) of 19 percent, with standard deviation of 0.023 mg/L and a mean of 0.121 mg/L for the 12-month period. The COV is defined as the standard deviation divided by the mean, and it is a measure of a system's reliability. The lower the COV, the less the data are spread and the higher the reliability. Variation is slightly higher on a weekly basis, with a COV of 41 percent. Overall, the facility is highly reliable at removing phosphorus. This is remarkable in comparison to many other facilities, which have reported poor reliability for biological phosphorus removal.

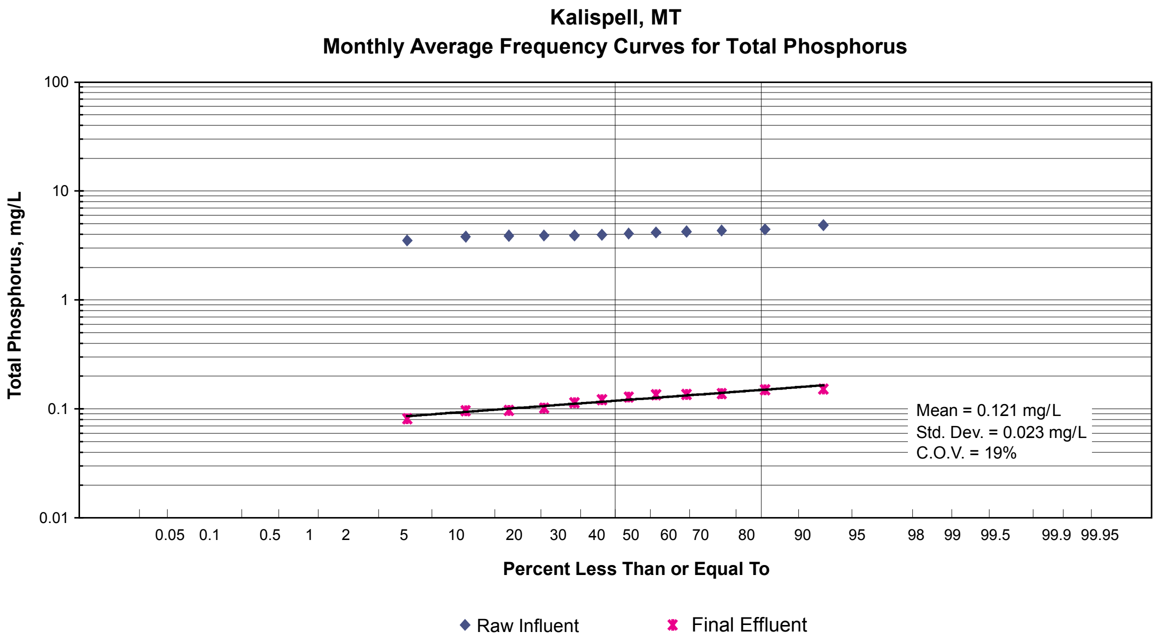

Figure 3. Monthly average frequency curves for TP.

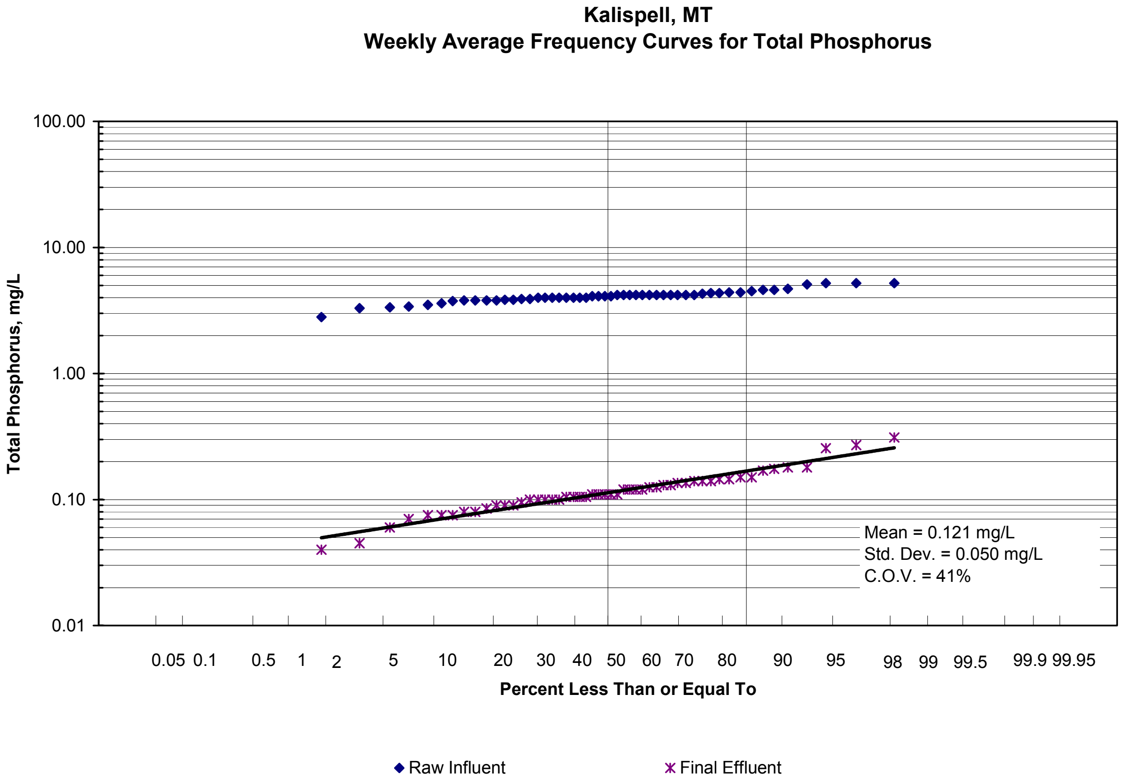

Figure 4. Weekly average frequency curves for TP.

Figure 5 presents the reliability plot for monthly average ammonia nitrogen. The facility reports only a monthly result for nitrogen compounds, which precludes generating a reliability plot for weekly data. For the period of July 2005 to June 2006, the plant routinely produced effluent ammonia nitrogen below a detection level of 0.07 mg/L. This is remarkable for a cold-region operation with an average water temperature of 8 degrees Celsius (°C) on cold days. The plant's successful operating strategy has been to maintain sufficient biomass during the winter, i.e., 2,700 parts per million (ppm) of mixed liquor suspended solids (MLSS) vs. 2,500 ppm in the summer. The higher biomass in winter allows the process to overcome the greatly slowed growth of nitrifiers under cold conditions.

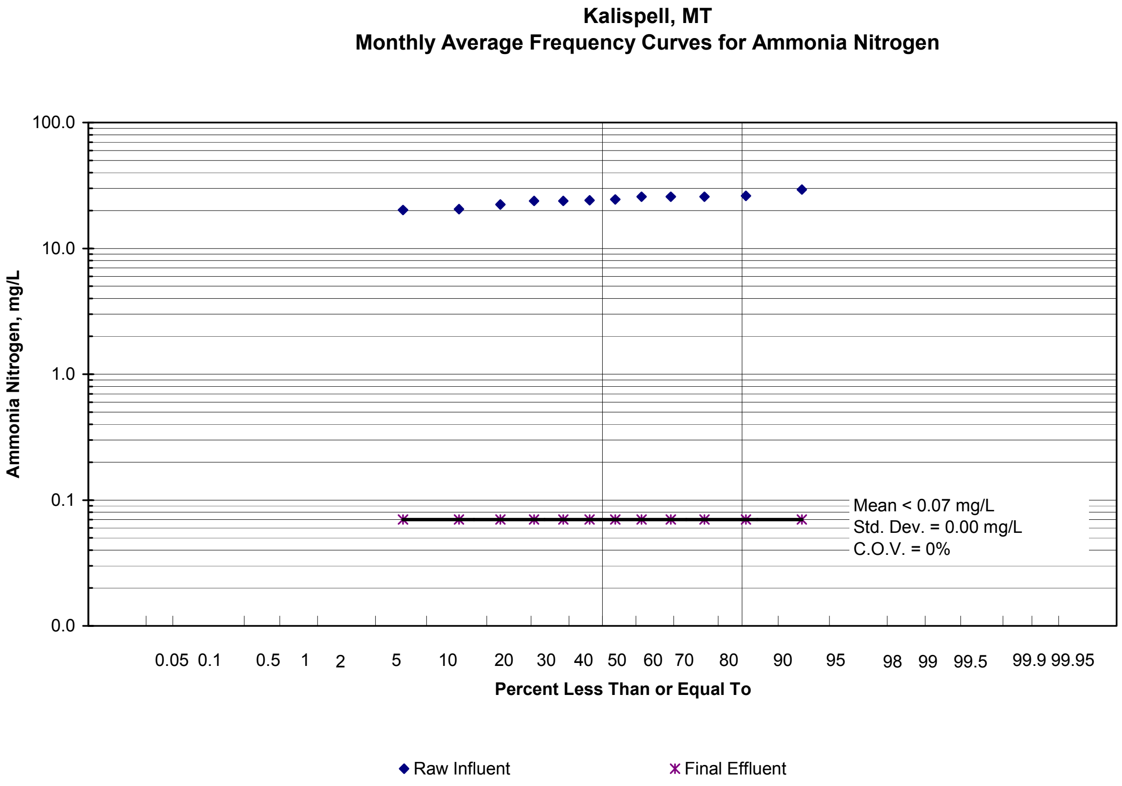

Figure 5. Monthly average frequency curves for ammonia nitrogen.

Figure 6 presents the reliability plot for monthly TN. The plant personnel set a design goal of 900 lb/day as TN (36 mg/L at 3 million gallons per day [MGD]), but this is not a permit limit. For the period of July 2005 to June 2006, the plant produced an effluent with an average TN of 10 mg/L, with more than 90 percent of that in the form of nitrate.

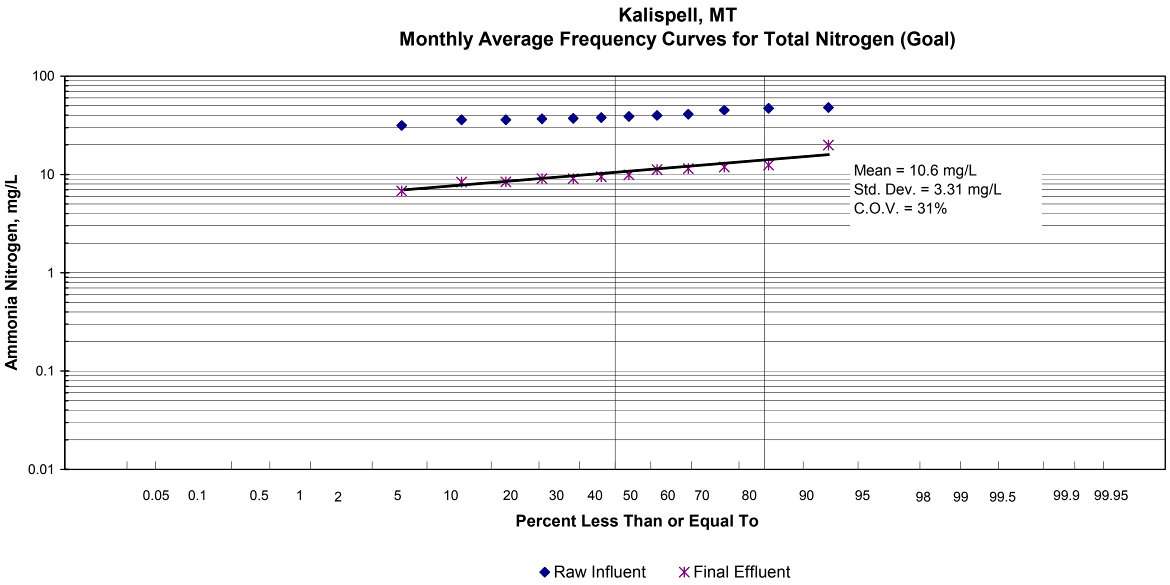

Figure 6. Monthly average frequency curves for TN (goal).

Reliability Factors

The plant has a permit limit for phosphorus of 1 ppm year-round monthly average; for ammonia nitrogen, it has a permit limit of 1.4 ppm monthly average to meet all stream requirements. However, the plant has an operational policy to achieve the maximum nutrient reduction without needing to add chemicals to precipitate phosphate or to support denitrification.

The key factor in the facility's success is generating sufficient volatile fatty acids (VFAs). The plant routinely meets its target of 18 mg/L VFAs at 20 °C and 13 mg/L VFAs at 13 °C in the anaerobic zones. This means that the VFA-to-total phosphorus (TP) ratio ranges seasonally between 1.5 and 6. The yearly average ratio is 3.5. The plant uses a two-stage fermenter to generate VFAs from primary sludge and produces around 200 mg/L VFAs in winter and 450 mg/L VFAs in summer under the sludge age of 4 to 5 days and an HRT of 7 to 21 hours. Unique design allows separate control of the SRT and HRT at this facility. Thickened fermented sludge is transferred to the anaerobic digesters, while the supernatant is pumped to the first anaerobic cell in the biological nutrient removal (BNR) system (Emrick and Abraham 2002; Natvik et al. 2003). The result is that the plant obtains effluent TP concentrations averaging 0.12 mg/L over the year with a low COV.

Another factor in the facility's success is that the plant personnel monitor each cell in the biological reactor for nutrient concentration, pH, and suspended solids and take actions as needed. Personnel do the monitoring by daily analyzing grab and composite samples rather than by using online sensors. The *hands-on* approach and daily attention to system performance prevent problems from becoming uncontrolled, while giving the operators a stake in the plant performance rather depending on the computer. Adjustments that can be made include solids wasting rate, recycle points, and which cells are aerobic or anoxic.

The flexibility in the process design is another valuable feature at Kalispell because the plant personnel can change the effective volumes of the anaerobic, anoxic, and aerobic zones by independently adjusting the conditions in each reactor cell as conditions warrant. The bioreactor is optimized for SRT and HRT at varying temperatures.

Another important operating practice is that of not maintaining sludge blankets in the secondary clarifiers (No Blanket Policy). This has helped the plant to achieve healthy biology with sufficient sludge age and excellent phosphorus removal because maintaining an inventory of sludge that has accumulated phosphorous maintains the chance that some of that phosphorous will eventually be released. In the summer the sludge age is maintained at between 8 and 10 days with an MLSS of 2,500 ppm. In winter the MLSS is increased to 2,700 ppm to ensure full nitrification under cold weather conditions.

Although this facility nitrified fully down to the detection limit (0.07 mg/L), the denitrification was not required and therefore was not practiced. The COV for ammonia nitrogen was 0 percent at the mean concentration of 0.07 mg/L as nitrogen. The COV was 31 percent at the mean concentration of TN of 10.6 mg/L.

Recycle loads were kept low at this facility. Secondary sludge was kept aerobic until dewatering, and the digester supernatant was kept at a minimum. The results were that the ortho-phosphorus returning to the headworks was measured at 6 percent of the influent TP load.

Wet-weather flows were managed through the equalization basin, which can store 12.5 percent of the influent flow. No special mode of operation was required at this facility.

Costs

Capital Costs

The plant was upgraded for BNR in 1992, when the system was set up as an 11-cell modified UCT with swing zones. The modifications for BNR were part of an overall upgrade program that cost a total of $13.5 million—$9.94 million in construction costs and $3.56 million in indirect costs. The elements involved in BNR that were included in the 1992 expansion are shown in Attachment 1. They included additional tanks, tank coatings, a supervisory control and data acquisition (SCADA) system, mixers, pumps, blowers, a fermenter, and a secondary sludge thickener. As shown in Attachment 1, these costs were attributed to removal of phosphorus, removal of nitrogen, or removal of non-nutrients, specifically biochemical oxygen demand (BOD). For units where the purpose could be fixed on one nutrient (e.g., a fermenter, which is only for phosphorus removal), the cost was attributed entirely to that nutrient. For the anoxic zone mixers, the cost was evenly divided between nitrogen and BOD removal because they are removed equally in those zones during denitrification. For the aeration zones and where units could not be specified for nutrients, the distribution was 12 percent for phosphorus, 48 percent for nitrogen, and 40 percent for BOD, which is the ratio at which those three removal processes take up oxygen on a molar basis during aeration.

The total of the construction costs for the BNR units was $4.2 million. Because the total indirect costs on the $9.9 million construction were $3.56 million, the indirect costs attributed to BNR were $1.51 million by ratio. These costs were allocated to phosphorus, nitrogen, and BOD removal using the 12/48/40 formula, resulting in $749,000 for phosphorus removal, $2.71 million for nitrogen removal, and $2.26 million for BOD removal, all in 1992 dollars.

These capital cost results were updated to 2007 dollars using the *Engineering News-Record*'s Construction Cost Index (ENR CCI). The ENR CCI, compiled by McGraw-Hill, provides a means of updating historical costs to account for inflation, thereby allowing comparison of costs on an equal basis. From a Web site provided by the U.S. Department of Agriculture, the ENR index for 1992 was 4,985, while the ENR index for May 2007 was 7,942 (USDA 2007). Multiplying the above results by the ratio 7,942/4,985 obtained the result of $1.19 million for phosphorus removal, $4.31 million for nitrogen removal, and $3.60 million for BOD removal in 2007 dollars.

These results were annualized using the interest rate formula for determining a set of annual payments for a present value, given an interest rate and payback period. For this and all other case studies for this document, a 6 percent interest rate and 20-year payback was assumed, resulting in a multiplication factor of 0.0872. The annualized capital cost for phosphorus removal was thus $101,500, while the annualized capital cost for nitrogen removal was $376,000. This annualized capital cost for nitrogen removal was used for later unit cost estimates for TN.

As shown in Attachment 1, the total capital charge for the BNR removal system was $5.7 million in 1992 dollars, which updated to $9.1 million in 2007 dollars. For this 3-MGD facility, the total capital cost for BNR removal was $3.03/gallon of treatment capacity.

Operation and Maintenance Costs

In all case studies prepared for this document, the O&M costs considered were for electricity, chemicals, and sludge disposal. Labor costs for operation and maintenance were specifically excluded for three reasons:

1. Labor costs are highly sensitive to local conditions, such as the prevailing wage rate, the relative strength of the local economy, the presence of unions, and other factors; thus, they would only confound comparison of the inherent cost of various technologies.
2. For most processes, the incremental extra labor involved in carrying out nutrient removal is recognized but not significant in view of the automatic controls and SCADA system that accompany most upgrades.
3. Most facilities were unable to break down which extra personnel were employed because of nutrient removal and related overtime costs, making labor cost development difficult.

The Kalispell plant uses an entirely biological process to achieve both nitrogen and phosphorus limits; therefore, the only significant operating cost is electrical use for mixers, pumps, and operating the fermenter. Attachment 2 shows a summary of the power use calculations. The power use attributed to phosphorus removal was 389,000 kilowatt-hours (kWh); using the average electrical rate of $0.045/kWh (which included all demand charges), the electrical cost for phosphorus removal was $17,500. The power usage attributed to nitrogen removal was 1,077,000 kWh, and at the average electrical rate, the electrical cost for nitrogen removal was $48,500.

Unit Costs for Nitrogen and Phosphorus Removal

In the evaluation period, the plant removed 35,700 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal was $0.49/lb, while the annualized unit capital cost for phosphorus removal was $2.84.

In the evaluation period, the plant removed 258,000 lb of TN. With the results above, the unit O&M cost for TN removal was $0.19/lb of TN, while the annualized unit capital cost for TN removal was $1.46.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the annualized unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal was $3.33/lb and the life-cycle cost for TN removal was $1.64/lb.

Assessment of magnitude of costs: The capital cost of $3.03/gpd capacity is relatively high, but the O&M costs are very low. One of the key factors is that chemicals are not used for nutrient removal, saving both those costs and costs that would be attributed to additional sludge generation.

Summary

The Kalispell Advanced WWTP has proven to successfully provide enhanced biological phosphorus removal in a cold-climate region of the United States. The reliability of the facility is good, with a mean effluent concentration of 0.12 mg/L as TP and a COV of 19 percent monthly average, or a COV of 41 percent weekly average. Ammonia nitrogen removal reliability is outstanding, with a mean concentration at or below the detection limit of 0.07 mg/L and a COV of 0 percent on a monthly average basis.

Reliability factors include a science-based control strategy, in-house generation of sufficient VFAs in the fermenter, and diligent monitoring and timely control of key process parameters by plant personnel. Removal costs for both phosphorus and nitrogen were shown to be reasonable, with O&M costs for both being largely driven by electricity usage and relatively low capital costs.

Acknowledgments

The authors of this report are grateful to Joni Emrick, water resource manager, and Curt Konecky of the Kalispell Advanced WWTP for their guidance and assistance in preparing this case study. This case study report would not have been possible without their prompt response with well-deserved pride in the facility and its operation. The authors also wish to thank the city of Kalispell for its participation.

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Attachment 1: Capital Costs

		%P	%N	%BOD	$P	$N	$BOD
Tanks	$1,300,000	12%	48%	40%	$156,000	$624,000	$520,000
Tank coats	$75,000	12%	48%	40%	$9,000	$36,000	$30,000
SCADA	$1,000,000	12%	48%	40%	$120,000	$480,000	$400,000
Mixers	$43,000	0%	50%	50%	$0	$21,500	$21,500
Ret/Sup pumps	$175,000	12%	48%	40%	$21,000	$84,000	$70,000
Blowers	$155,000	0%	50%	50%	$0	$77,500	$77,500
Fermenter	$45,000	100%	0%	0%	$45,000	$0	$0
Thickener	$35,000	100%	0%	0%	$35,000	$0	$0
Primary sludge pump	$80,000	10%	50%	40%	$8,000	$40,000	$32,000
Piping	$500,000	12%	48%	40%	$60,000	$240,000	$200,000
Site work	$800,000	12%	48%	40%	$96,000	$384,000	$320,000
Total	$4,208,000				$550,000	$1,987,000	$1,671,000
Indirects	$1,505,526	12%	48%	40%	$180,663	$722,653	$602,211
Total capital	$5,713,526				$730,663	$2,709,653	$2,273,211
Updated to 2007	$9,102,673				$1,164,078	$4,316,963	$3,621,633
Annualized					$101,508	$376,439	$315,806
Updating factors							
1992 ENR CCI	4,985						
May 2007 ENR CCI	7,942						
A/P (6%, 20 years)	0.0872						

Attachment 2: Electrical Costs

Horsepower	Volts	Amps	VA	Number	kW Power draw	hours/day	kWh draw/day	kWh draw/year	%P	%N	For P	For N
Mixers												
3	460	4	1,840	5	9.2	24	220.8	80,592	12%	48%	9,671.04	38,684.16
7.5	460	10	4,600	4	18.4	24	441.6	161,184	12%	48%	19,342.08	77,368.32
Ret Pumps												
10	460	5.82	2,677.2	1	2.6772	24	64.2528	23,452.27	12%	48%	2,814.273	11,257.09
4	460	6.7	3,082	1	3.082	24	73.968	26,998.32	12%	48%	3,239.798	12,959.19
Blowers												
200	460	220	101,200	2	202.4	24	4,857.6	1,773,024	0%	50%	0	886,512
Super Pumps												
7.5	460	9.7	4,462	2	8.924	24	214.176	78,174.24	12%	48%	9,380.909	37,523.64
Fermenter												
5	460	6.8	3,128	2	6.256	24	150.144	54,802.56	100%	0%	54,802.56	0
15	460	27	12,420	2	24.84	24	596.16	217,598.4	100%	0%	217,598.4	0
10	460	14	6,440	1	6.44	24	154.56	56,414.4	100%	0%	56,414.4	0
Gravity Thickener												
2	460	3.1	1,426	1	1.426	24	34.224	12,491.76	100%	0%	12,491.76	0
Primary Clarifier Sludge Pump												
5	460	6.8	3,128	1	3.128	24	75.072	27,401.28	12%	48%	3,288.154	13,152.61
							kWh/yr	2,512,133			389,043.4	1,077,457
							Rate	0.045	$/kWh		P	N
							Totals	**113,046**		**$/yr**	**17,506.95**	**48,485.57**

Clark County Water Reclamation Facility
Las Vegas, Nevada
Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

The Clark County Water Reclamation Facility (WRF) is in Las Vegas, Nevada. This facility was selected as a case study because of the anoxic/oxic (A/O) process for biological phosphorus removal with alum feed.

Originally commissioned in 1956, the facility was enhanced with biological nutrient removal (BNR) in 1995 during an 88-million gallon per day (MGD) expansion. The plant has obtained a very high level of phosphorus removal following a series of facility upgrades.

With the expansion, the facility essentially operates as two plants—the Advanced Waste Treatment Plant (AWT) and the Central Plant (CP)—with separate discharges available. The expansion allowed the plant to gain nitrification capabilities for the entire plant flow, in both the CP and the AWT. Although the facility initially used and still uses chemical treatment to meet standards, it has also implemented the A/O process to provide biological phosphorus removal. The facility is designed for an average flow of 100 MGD and averaged 95 MGD during the 2006 calendar year.

The relevant National Pollutant Discharge Elimination System (NPDES) permit limits are listed in Table 1.

Table 1. Clark County WRF NPDES permit limits

Parameter	30-day avg. (mg/L)	7-day avg. (mg/L)	30-day avg. (lb/day)	Daily wasteload allocation (lb/day)
BOD	30	45	37,530	--
TSS	30	45	37,530	--
TP	--	--	--	173
Total NH_4-N	--	--	--	502

Notes:
BOD = biochemical oxygen demand
NH_4-N = ammonia measured as nitrogen
P = phosphorus
TSS = total suspended solids

The wasteload allocation is an arrangement in which the Nevada Division of Environmental Protection set an overall load on the Las Vegas Wash from Clark County, the city of Las Vegas, and Henderson, Nevada. The allocations for Clark County translate into 0.21 milligrams per liter (mg/L) total phosphorus (TP) and 0.6 mg/L for ammonia nitrogen at 100 MGD.

Basis of Design and Flow Schematic

Primary settling tanks: 818 gallons per day per square foot (gpd/ft^2) at annual average flow and 1,309 gpd/ft^2 at peak hour

Activated sludge: nine basins

Hydraulic capacity per basin	10 MGD
Total volume per basin	2.13 MG
Hydraulic retention time	5.1 hours
Sludge age	5–9 days

Secondary clarifier: 710 gpd/ft^2 at annual average flow

A flow sheet for the CP is presented in Figure 1 for the entire facility. The main difference between the AWT and the CP is that the AWT employs tertiary clarifiers in advance of the tertiary filters, as shown in Figure 2. In both plants, influent is treated in the primary settling tanks with ferric chloride added as enhancement, then through A/O biological reactors. The A/O process provides biological phosphorus removal and nitrification, along with some degree of denitrification. From there, the wastewater is dosed with alum for additional phosphorus removal and then treated in a tertiary clarifier/filter combination in the AWT or in just a tertiary filter in the CP. When the clarifiers were first installed in the 1980s, filter technology was such that they needed protection from high solids that would make operation and maintenance (O&M) difficult; the CP uses an air-water, scour-backwash system so that such protection is not vital to continued good operation. The effluent is filtered and disinfected by ultraviolet (UV) radiation and then either sent to reclaimed water customers or discharged to the Las Vegas Wash and the Lake Meade Wetlands.

The secondary sludge is thickened by dissolved air flotation (DAF). The primary sludge is thickened to 5 percent solids in the settling tanks and then sent to the same holding tank with the thickened secondary sludge. They are dewatered together by belt filter press for landfilling.

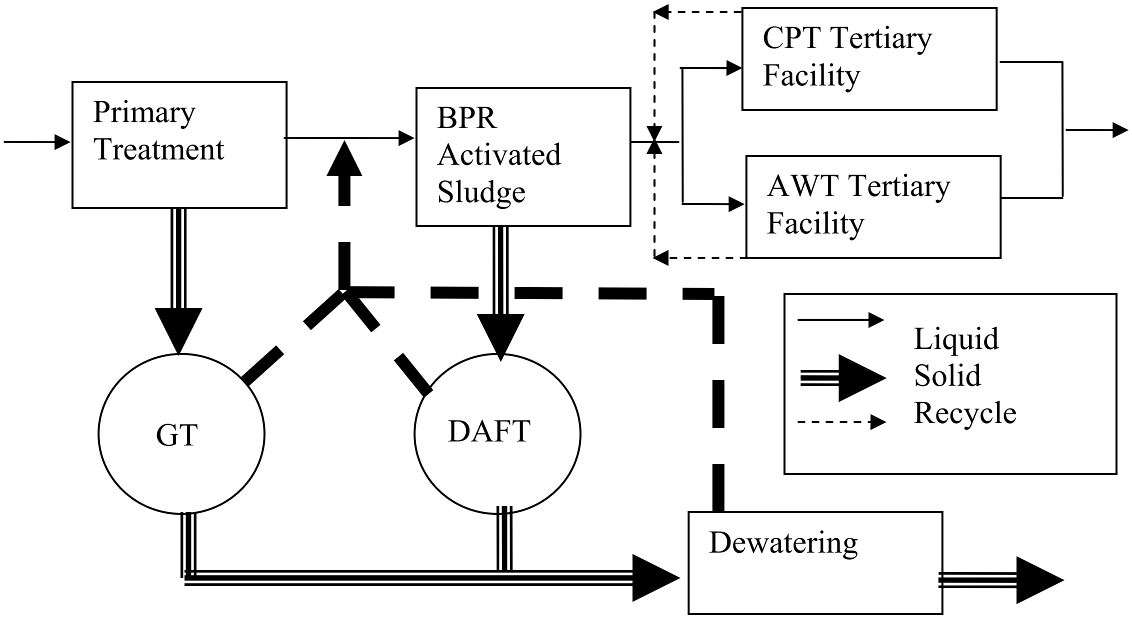

Figure 1. Plant flow schematic.

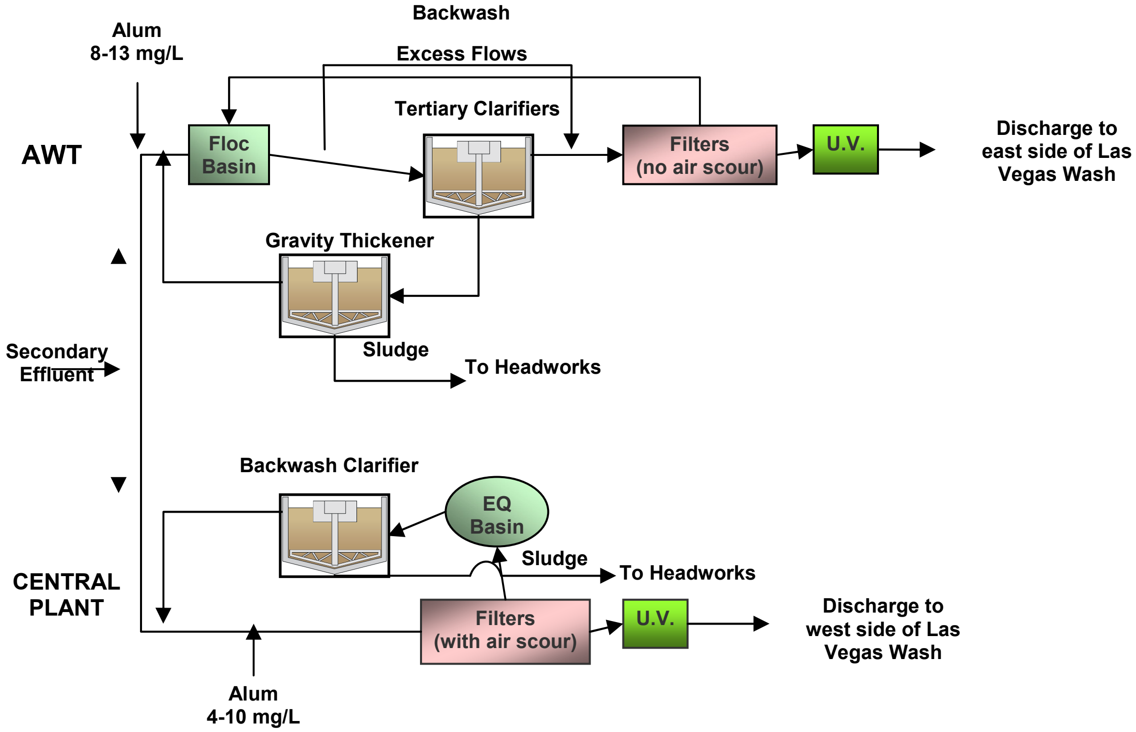

Figure 2. Tertiary processes.

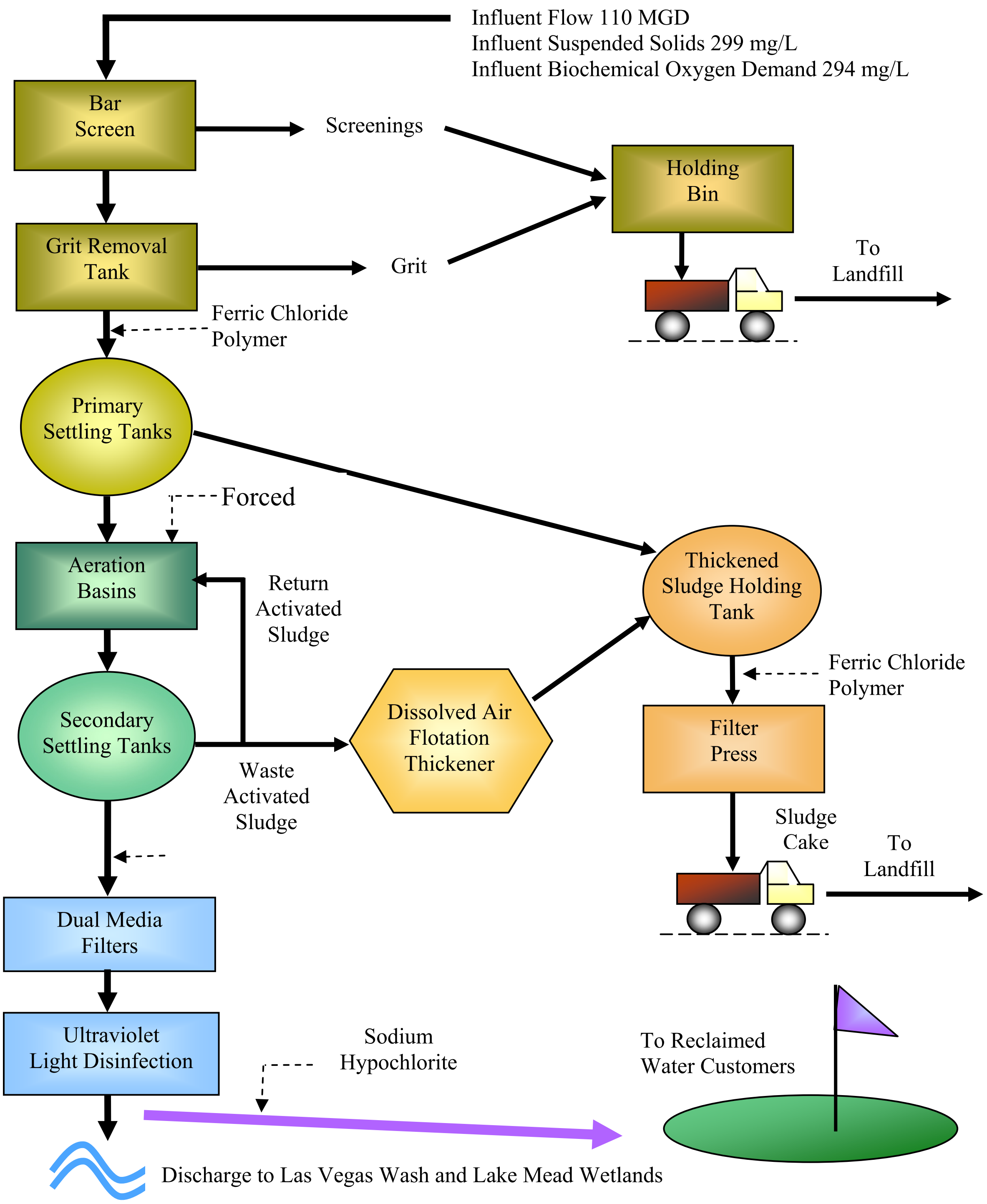

Figure 3. Clark County WRF CP flowsheet schematic.

Plant Data

Table 2 presents average plant data for the 2006 calendar year. The data show outstanding removal of nutrients, biochemical oxygen demand (BOD), and suspended solids. The facility easily meets all of its permit limits.

Table 2. 2006 average CP water quality data

Parameter	Average	Max month	Max month vs. avg.	Max week	Sample method/ frequency
Flow (MGD)	98	101.4	3.5%	102.3	--
Influent TP (mg/L)	5.8	7.0	20%	7.5	Daily/composite
Effluent TP (mg/L)	0.1	0.17	73%	0.41	Daily/composite
Influent BOD (mg/L)	357	390	9%	445	Daily/composite
Effluent BOD (mg/L)	< 2	4.75	137%	7	Daily/composite
Influent TSS (mg/L)	366	413	13%	456	Daily/composite
Effluent TSS (mg/L)	< 5	10	100%	21	Daily/composite
Influent NH_4-N (mg/L)	26.8	28.8	7%	30	Daily/composite
Effluent NH_4-N (mg/L)	0.08	0.31	300%	1.22	Daily/composite
Influent TKN (mg/L)	46	53	14%	75	Daily/composite
Effluent TKN (mg/L)	0.69	1.02	47%	2.3	Daily/composite
Influent NO_3/NO_2 (mg/L)	0.18	0.46	155%	0.8	Daily/composite
Effluent NO_3/NO_2 (mg/L)	15.3	16.4	7%	16.5	Daily/composite
Influent TN (mg/L)	30.3	34.5	14%	37.6	--
Effluent TN (mg/L)	15.2	16.6	7%	16.7	--

Notes:
BOD = biochemical oxygen demand
NH_4-N = ammonia measured as nitrogen
NO_3/NO_2 = nitrate + nitrite
TKN = total Kjeldahl nitrogen
TN = total nitrogen
TP = total phosphorus
TSS = total suspended solids

Table 3 presents plant monthly average plant process parameters.

Table 3. CP monthly average plant process parameters

Month	MLSS (mg/L)	Sludge age (d)	HRT (hr)	Temperature (°C)
Jan 2006	2,902	9	5.43	20
Feb 2006	3,422	9	5.45	20
Mar 2006	3,684	8	5.31	24
Apr 2006	3,732	7	5.10	26
May 2006	3,147	6	5.06	28
June 2006	3,499	5	5.82	29
July 2006	3,166	5	5.73	29
Aug 2006	3,057	5	5.80	29
Sept 2006	2,425	6	6.32	28
Oct 2006	2,441	7	6.31	26
Nov 2006	2,760	8	6.42	24
Dec 2006	2,535	8	6.49	20

Notes:
HRT = hydraulic retention time
MLSS = mixed liquor suspended solids

Performance Data

Figures 4 and 5 present reliability plots for weekly average and monthly average TP. The plant operation provides outstanding performance in TP removal: the average effluent concentration is under 0.1 mg/L and the coefficient of variation (COV) is low at 30 percent. This means that the data have a low standard deviation relative to the mean and, therefore, that the plant will routinely produce effluent with TP below 0.2 mg/L through the course of the year.

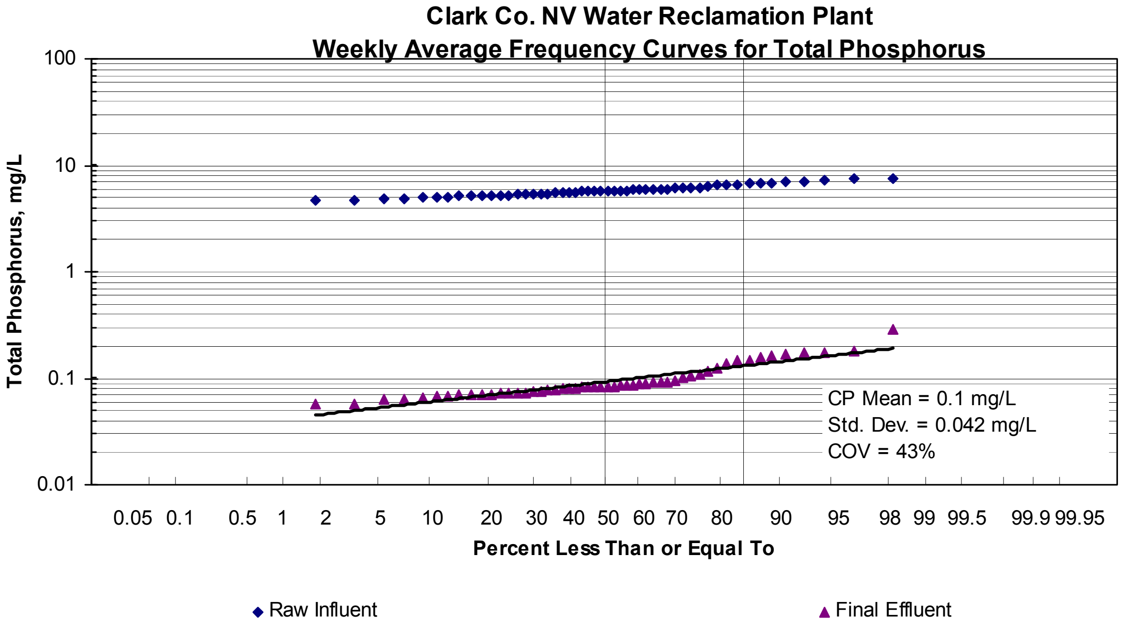

Figure 4. Weekly average frequency curves for TP.

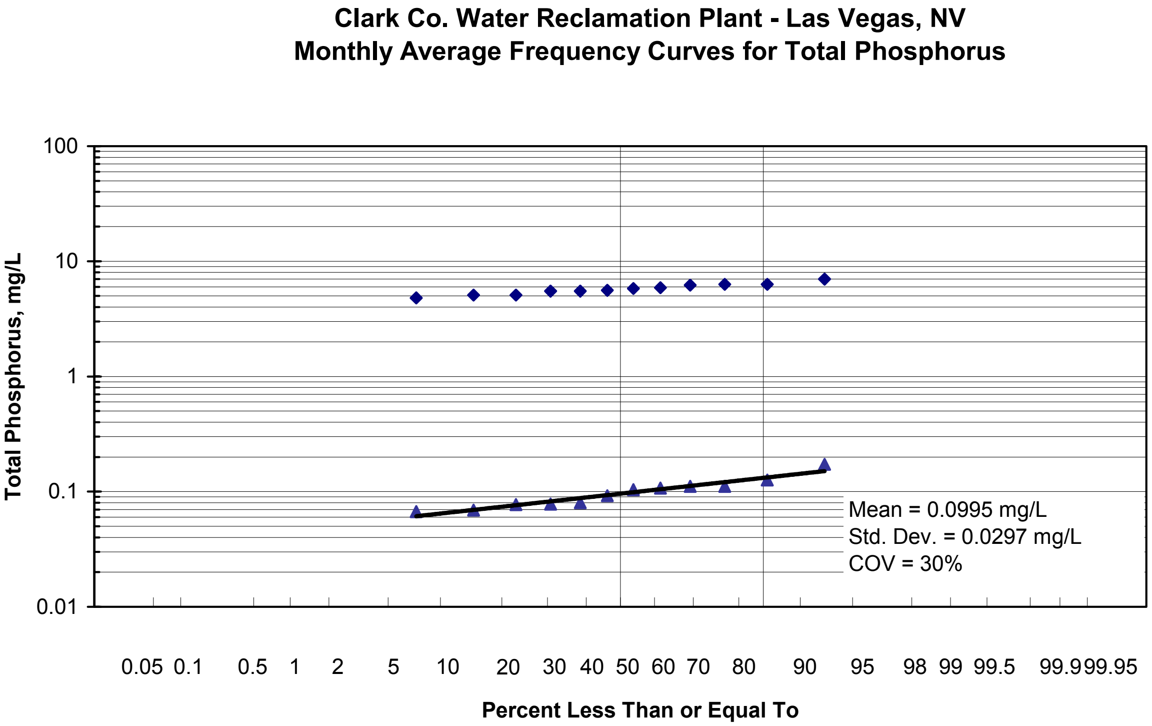

Figure 5. Monthly average frequency curves for TP.

Figures 6 and 7 present reliability plots for the weekly average and monthly average total nitrogen (TN) for the facility. TN removal is not required under the permit, and therefore it is limited. The effluent TN averages 15.2 mg/L with a standard deviation of 0.6 mg/L.

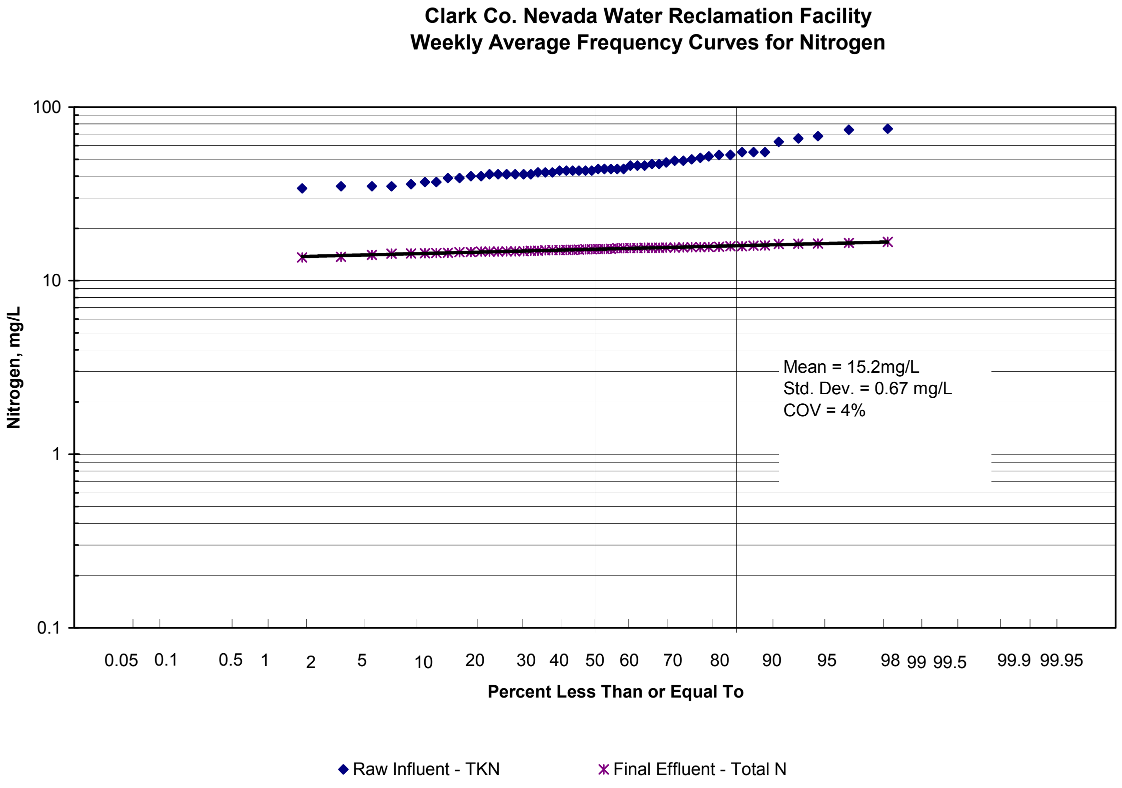

Figure 6. Weekly average frequency curves for nitrogen.

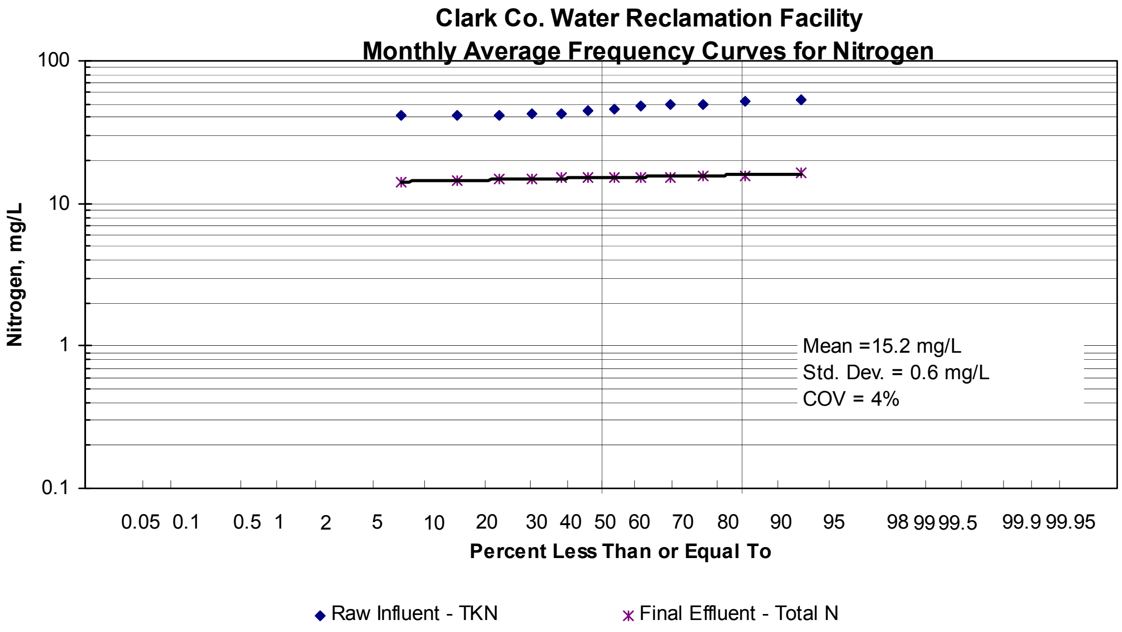

Figure 7. Monthly average frequency curves for nitrogen.

Figures 8 and 9 present reliability plots for weekly average and monthly average ammonia nitrogen for the plant. Ammonia is routinely removed to near the detection level in the plant, with a mean of 0.05 mg/L and a very low COV of 22 percent.

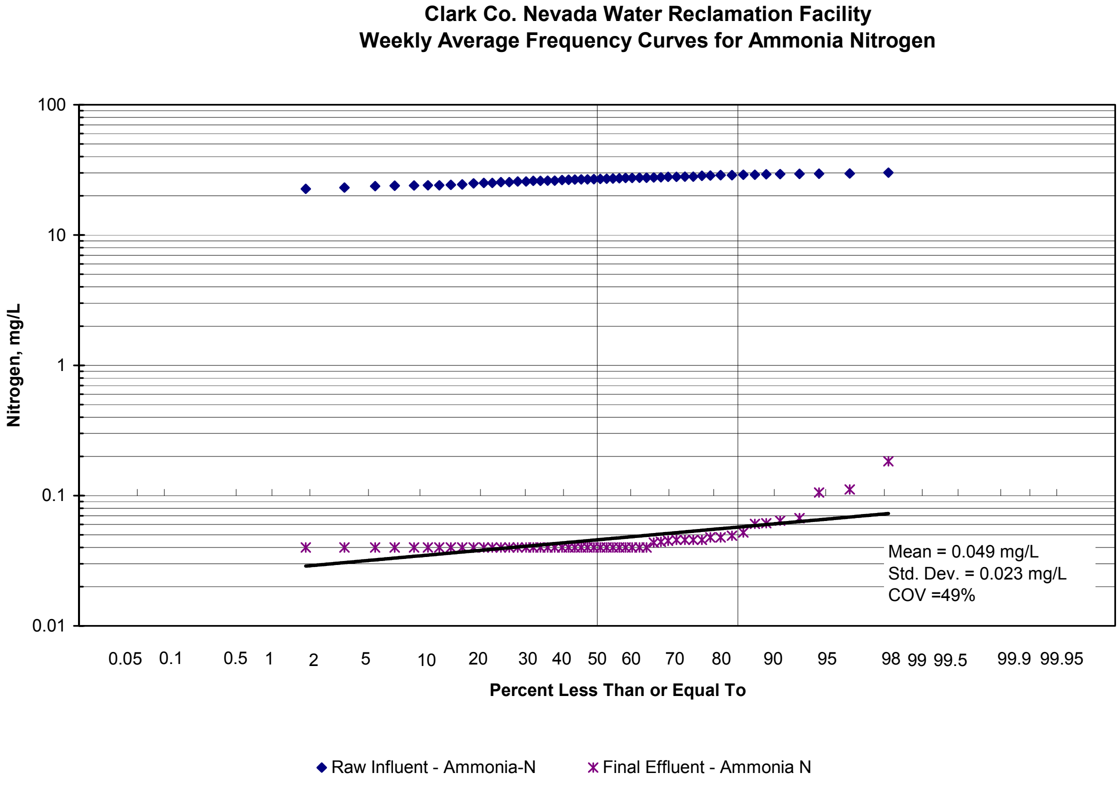

Figure 8. Weekly average frequency curves for ammonia nitrogen.

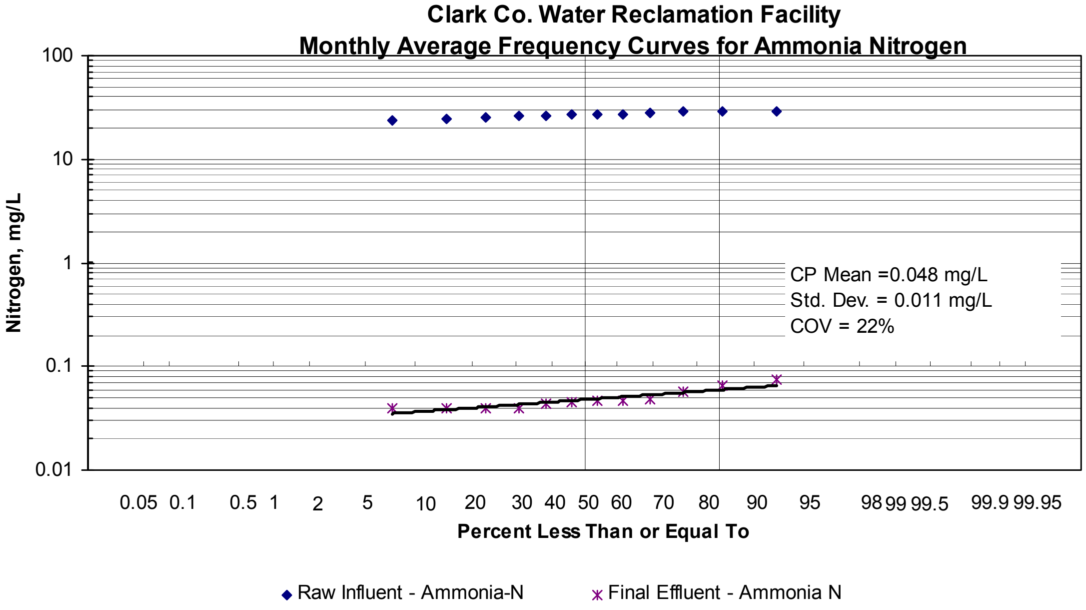

Figure 9. Monthly average frequency curves for ammonia nitrogen.

Reliability Factors

Several factors have contributed to efficient and reliable operation at this facility. The effluent concentration was low at 0.09 mg/L in TP with a COV of 30 percent and 0.05 mg/L in ammonia nitrogen with a COV of 22 percent.

One key is the wastewater characteristics and in-plant generation of volatile fatty acids (VFAs). The BOD-to-TP ratio averaged 29.8 for the year and ranged from an average 26.5 to 34.2 monthly. Furthermore, this facility generated additional VFAs by operating primary settling tanks as fermenters. Typical operating parameters included thickening the primary sludge to 5 percent total solids, thereby generating enough VFAs to maintain 35 to 45 mg/L of VFA in the primary effluent. Thickening primary sludge to 6 percent total solids was found excessive and detrimental to both odor-control and clarification purposes.

The biological process was originally a conventional process, which was later converted to an A/O process by adding aeration controls to ensure sufficient dissolved oxygen (DO) in the aerobic zones. The DO set point is 2.4 mg/L to meet an instantaneous minimum DO of 2.0 mg/L. The optimal sludge age ranged from 5 days in summer at 29 degrees Celsius (°C) to 9 days in winter at 20 °C. The average secondary effluent concentration showed an average of 0.7 mg/L as TP, 0.1 mg/L ammonia nitrogen, and 15 mg/L in TN, with a return activated sludge (RAS) flow ranging from 45 to 60 percent. The clarifiers are operated with a minimal blanket (less than 6 inches) to prevent secondary release of phosphorus. Secondary release of phosphorus is of concern at this plant because of the generally high temperatures increasing biological activity.

Another factor is the successful polishing of the biological process effluent for phosphorus by the tertiary clarifiers and filters. The AWT has a tertiary clarifier ahead of tertiary filters and performs better than the CP when the biological phosphorus removal process is upset and carries elevated levels of suspended solids. The tertiary clarifier acts as an added line of defense for the filters and maintains steady effluent quality ahead of the filters. At the AWT, alum addition can go up to 15–16 mg/L without a having an adverse effect on the filters. The CP, however, does not have a tertiary clarifier, and the alum dosage is limited to 10–12 mg/L before the filters become blinded by solids. Note that filters at the CP have an air-water backwash capability and therefore work well under these operating conditions.

Another key to successful removal of phosphorus is having multiple chemical feeding points. Ferric chloride is fed to the primary settling tanks with the primary purpose of removing suspended solids and a resulting side benefit of removing some phosphorus. The dosage of ferric chloride averages 10–12 mg/L. Alum is added as described above to polish residual phosphorus ahead of the tertiary filters.

Another key to successful phosphorus removal is minimal recycle of in-plant phosphorus loads. Waste activated sludge (WAS) is thickened in a dissolved air floatation (DAF) process, and the combined primary sludge (0.7 MGD) and WAS sludge (1.15 MGD) are dewatered daily at the belt filter press with ferric chloride and polymer addition. This operation minimizes the release of phosphorus and prevents odor generation. The key operational activity here is the daily dewatering of all sludge. Reduction in odors is also aided by processing the sludge daily, which is accomplished by plant personnel working two 10-hour shifts and processing all sludge generated at the plant. This practice ensures a minimal amount of odor generation at the plant and the minimum recycle of phosphorus loadings back to the treatment processes. The TP in the filtrate from dewatering ranges between 100 and 300 mg/L. The TP in the recycle flows is in the range of 20 to 25 percent of the influent total.

The final line of defense is the tertiary filters. They were designed to operate at 5 gpm/ft^2 during dry-weather peak flows and have performed well. The maintenance dosage of alum is fed into tertiary filters to prevent secondary release from biological solids. They average 6 mg/L at the AWT and 4 mg/L at the CP. The long-term average soluble phosphorus leaving the filters is less than 0.02 mg/L.

A benefit of having biological phosphorus removal followed by chemical polishing is reduction in chemical sludge. Over the years, the plant has observed a decrease in total sludge production. In 1997 the average sludge production was approximately 600 wet tons per day. In 2007 even with increased flows, the sludge production is approximately 400 wet tons per day.

Another key in the successful operation of the plant was automating the process monitoring and controls. Two distinct functions are automated at this plant. One is that the decisions on WAS from nine separate trains are made and carried out by a program developed in-house using a mixed liquor suspended solids (MLSS) probe. The other is automatic blower control in the aerobic zones. The head section of the aerobic zone receives the maximum supply of air, while the latter section of the zone is controlled by a program with a set point of 2.4 mg/L DO using multiple probes.

The blowers are a key part of the process and require redundancy. The operating philosophy is to provide a minimum of 0.5 mg/L DO at all times, even during the peak hot period of the day. The plant experienced a DO deficit during a week of air temperatures at 113 degrees Fahrenheit (°F) (45 °C), which was detrimental to the biological treatment process.

Another key is good redundancy, achieved by running nine separate treatment processes in parallel. If one train experiences an upset condition, operators can supply good seed MLSS from one of the other trains.

Alternative Processes Considered

Because the plant is almost at capacity (100 MGD versus 110 MGD), expansion plans are being pursued. For the AWT, a pilot program is underway for membrane filtration of secondary effluent. Three different membranes are being evaluated concurrently. If the evaluations are successful, the membrane filter could replace both the tertiary clarifier and the dual media filters.

Costs

Capital Costs

The plant has undergone a number of upgrades and renovations since the original commissioning of the AWT in 1982. Those total costs were updated to 2007 dollars using the *Engineering News-Record* Construction Cost Index for construction costs and the Consumer Price Index for the applicable years (USDA 2007). The resulting capital costs, the attributed percentages for phosphorus and nitrogen removal, and the resulting total capital costs are shown in Table 4.

Table 4. Upgrade capital costs and resulting phosphorus and nitrogen removal

Capital	Year	Amount	Updated cost	%P	%N	P removal	N removal
AWT Des	1982	$2,800,000	$5,956,103	50%	0%	$2,978,051	$0
AWT Const	1982	$28,000,000	$58,137,516	50%	0%	$29,068,758	$0
CP Expan Design	1994	$2,000,000	$2,770,875	12%	48%	$332,505	$1,330,020
CP Expan Const	1994	$29,000,000	$42,588,388	12%	48%	$5,110,607	$20,442,426
CP Filters Design	2002	$4,200,000	$4,794,056	50%	0%	$2,397,028	$0
CP Filters Const	2002	$27,600,000	$33,526,950	50%	0%	$16,763,475	$0
Central Plant S. Sec. Design	2003	$3,790,000	$4,230,603	12%	48%	$507,672	$2,030,689
Central Plant S. Sec. Const	2003	$39,,304,293	$46,625,048	12%	48%	$5,595,006	$22,380,023
Central Plant S. Sec. Design	2005	$1,901,098	$1,998,417	12%	48%	$239,810	$959,240
Central Plant S. Sec. Const	2005	$19,218,993	$20,499,227	12%	48%	$2,459,907	$9,839,629
TOTAL		$157,814,384	$221,000,000	--	--	$65,452,819	$56,982,027

The capital expenditure in 2007 dollars that could be attributed to phosphorus removal was $65.4 million. The annualized capital charge (20 years at 6 percent) was $5.71 million for phosphorus removal.

The capital expenditure in 2007 dollars that could be attributed to TN removal was $57 million. The annualized capital charge (20 years at 6 percent) was $4.97 million for TN removal. This same expenditure could be attributed to ammonia nitrogen removal.

The total capital attributed to BNR in 2007 dollars was $221 million. For the 110-MGD facility, the capital expenditure per gallon of BNR treatment capacity was $2.01.

Operation and Maintenance Costs

The Clark County plant uses a combination of biological and chemical phosphorus removal to achieve the limit. This means that costs for phosphorus removal are distributed among primary treatment (adding ferric chloride), secondary treatment (aeration basins, mixers, and pumps), tertiary treatment (chemical addition and filtration), solids dewatering, and laboratory testing. Costs for each of those components of wastewater treatment are shown in Table 5, with the percentages of the costs that were attributed to TP and TN removal and the final values.

Table 5. Component costs and resulting phosphorus and nitrogen removal

Component	Total op. costs	% for P	% for N	P O&M	N O&M
Primary	$1,877,685	12%	48%	$225,322	$901,289
Secondary	$5,829,302	12%	48%	$699,516	$2,798,065
Tertiary	$3,967,135	12%	0%	$476,056	$0
Solids dewatering	$3,957,135	50%	0%	$1,450,019	$0
Lab	$1,529,827	10%	10%	$152,983	$152,983
Other	$3,875,144	0%	0%	$0	$ 0
TOTAL	$19,979,131	--	--	$3,003,896	$3,852,337

Unit Costs for Nitrogen and Phosphorus Removal

In 2006 the plant removed 1,663,000 lb of phosphorus. With the results shown in Tables 3 and 4, the unit O&M cost for phosphorus removal is $1.81/lb, and the unit capital cost is $3.43/lb of phosphorus removed.

In 2006 the plant removed 8,994,000 lb of nitrogen. With the results shown in Tables 3 and 4, the unit O&M cost for nitrogen removal is $0.43/lb and the capital cost is $0.55/lb of TN removed.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal is $5.24/lb phosphorus removed and while the life-cycle cost for nitrogen removal is $0.98/lb nitrogen removed.

Assessment of Magnitude of Costs and Main Factors

The life-cycle costs for phosphorus removal and full nitrification are on the high side, for achieving an extremely low level of phosphorus and ammonia nitrogen by upgrading existing facilities.

Discussion

Reliability factors: Three major factors contribute to a reliable performance in phosphorus removal and nitrification: (1) multiple chemical feeds to the system, (2) good biological phosphorus removal with in-plant VFA generation and full nitrification, and (3) good tertiary filters in suspended solids removal. This combination of chemical, biological, and physical processes in series provides a reliable operation with exceptionally low concentrations of phosphorus at 0.09 mg/L with a low COV of 30 percent, while the ammonia nitrogen concentration is at 0.05 mg/L with an even lower COV of 22 percent average monthly.

Cost factors: This plant is an example of exceeding the original design capacity with retrofit upgrades, which results in significant cost savings. The capital cost for phosphorus removal and complete nitrification is estimated to be low at $2.01/gpd capacity. The unit costs for capital and O&M were $5.43/lb of phosphorus removed and $1.38/lb of nitrogen removed. The unit costs for O&M were $1.84/lb of phosphorus removed and $0.51/lb of nitrogen removed.

Summary

The Clark County plant operation has been successful in reducing effluent phosphorus to the limit of technologies at the existing plant using a combination of biological and chemical treatment processes in series with good reliability. The plant is almost at capacity and yet has produced effluent far below the discharge limits. The mean TP concentration was 0.099 mg/L for the year with a COV of less than 30 percent, at either the AWT or CP. The technique of using several different technologies in series to achieve the treatment objective works, especially when operation is done with computer control and the system has been designed with a reasonable amount of *robustness* to allow for repairs and routine maintenance. The instrumentation technician on staff is a unique and valuable member of the team at this facility. The costs of operation are also reasonable: life-cycle costs are $5.24/lb and $0.98/lb for phosphorus and nitrogen removal, respectively.

Acknowledgments

The authors are grateful for the significant assistance and guidance provided by Dr. Douglas Drury, deputy general manager, and Danielle Fife at the Clark County WRF. This case study

would not have been possible without their prompt response with well-deserved pride in their facility and its operation. EPA thanks Clark County for participating in this case study.

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Kelowna Wastewater Treatment Plant
Kelowna, British Columbia, Canada

Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

The Kelowna Wastewater Treatment Plant (WWTP) is in the province of British Columbia in western Canada. This plant was selected as a case study because of its cold-weather biological nutrient removal (BNR) with a five-stage Bardenpho process, which has been retrofitted into a new, three-stage Westbank process.

A BNR process, as depicted in Figure 1, was commissioned in 1982 and was operated successfully through the 1980s. Optimization was ongoing, and an understanding of the BNR removal mechanisms and pathways was developed, tested, and documented in Kelowna and through other worldwide research programs.

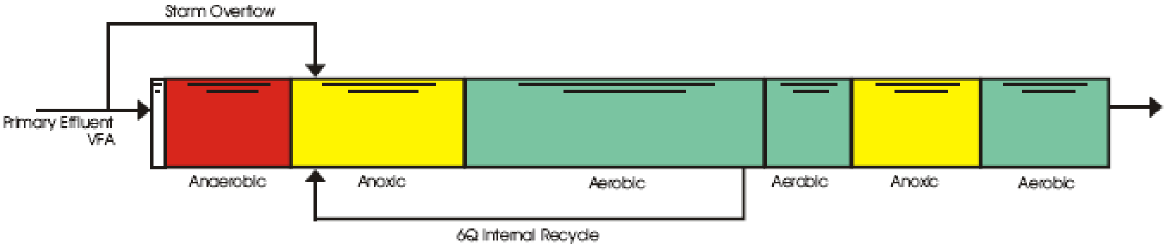

Figure 1. Kelowna five-stage Bardenpho process.

The Canadian Ministry of Environment (MOE) permit requirements, shown in Table 1, include biochemical oxygen demand (BOD_5)-total, total suspended solids (TSS), total nitrogen (TN), and total phosphorus (TP) limits. The plant's overall performance is shown in Table 2.

Table 1. Permit requirements for effluent quality

MOE permit requirements	Daily limits (mg/L)
BOD_5-total	8
TSS	7
TN	6
TP Maximum 99th percentile 90th percentile Annual average (added in 1988)	 2.0 1.5 1.0 0.25

Table 2. Influent and effluent averages

Parameter (mg/L unless stated)	Average	Maximum month	Max month vs. avg.	Maximum week	Sample method/frequency
Flow (MGD)	8.5	8.8	3.4%	8.9	--
Influent TP	6.0	7.4	23%	9.1	Composite/weekly
Effluent TP	0.14	0.20	42%	0.25	Composite/weekly
Influent COD	626	747	19%	910	Composite/weekly
Effluent COD	32	36	10%	38	Composite/weekly
Effluent BOD	2.5	3.8	48%	5.7	Composite/weekly
Influent TSS	389	472	21%	532	Composite/weekly
Effluent TSS	1.2	1.6	42%	2.4	Composite/weekly
Influent NH_4-N	21.3	23.1	8.3%	27.6	Grab/monthly
Effluent NH_4-N	0.57	1.0	76%	1.13	Grab/monthly
Influent TKN	28.8	33	14%	38.4	Grab/monthly
Effluent TKN	2.0	2.98	49%	3.5	Grab/monthly
Influent TN	28.8	33	14%	38.4	Grab/monthly
Effluent TN	4.38	4.9	12%	5.84	Grab/monthly

Notes:
BOD = biochemical oxygen demand
COD = chemical oxygen demand
Max month vs. average = (max month – average) / average x 100
MGD = million gallons per day
NH_4-N = ammonia measured as nitrogen
TN = total nitrogen
TP = total phosphorus
TSS = total suspended solids

Treatment Processes

As the load on the facility increased, it became clear that the five-stage process with a 22-hour hydraulic retention time (HRT) design far exceeded the HRT necessary to meet effluent discharge requirements for both TP (0.25 milligrams per liter [mg/L]) and TN (6.0 mg/L). Process developments led to implementing a *high-rate* BNR process that was initially tested at the Kelowna facility. The first full-scale implementation was at the Westbank WWTP 20 miles southwest of the Kelowna plant. Details of the basis for plant design are provided in Attachment 1.

Figure 2 depicts a shorter HRT process, and in 1994 the Kelowna facility was retrofitted in this mode of operation. In effect, the last two stages (anoxic and aerobic) were bypassed and made redundant. Later, the bypassed modules were retrofitted as two additional, smaller Westbank-type modules.

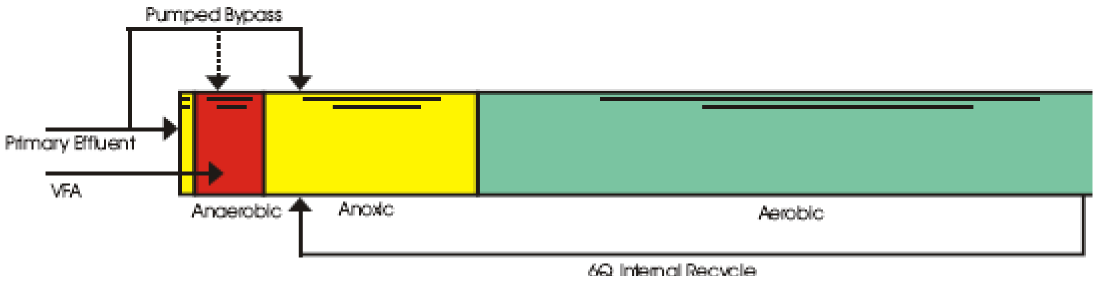

Figure 2. The Westbank three-stage process.

The Kelowna WWTP layout, as depicted in Figure 3, was implemented with the following process elements:

The liquid train includes

- Screening
- Grit removal
- Primary sedimentation
- Three-stage Westbank BNR configuration
- Secondary clarifiers
- Dual media filters
- UV disinfection
- Flow and load equalization

The solids train includes

- Primary sludge fermenter
- Air flotation for waste activated sludge (WAS) thickening
- Centrifuge
- Hauling to compost facility

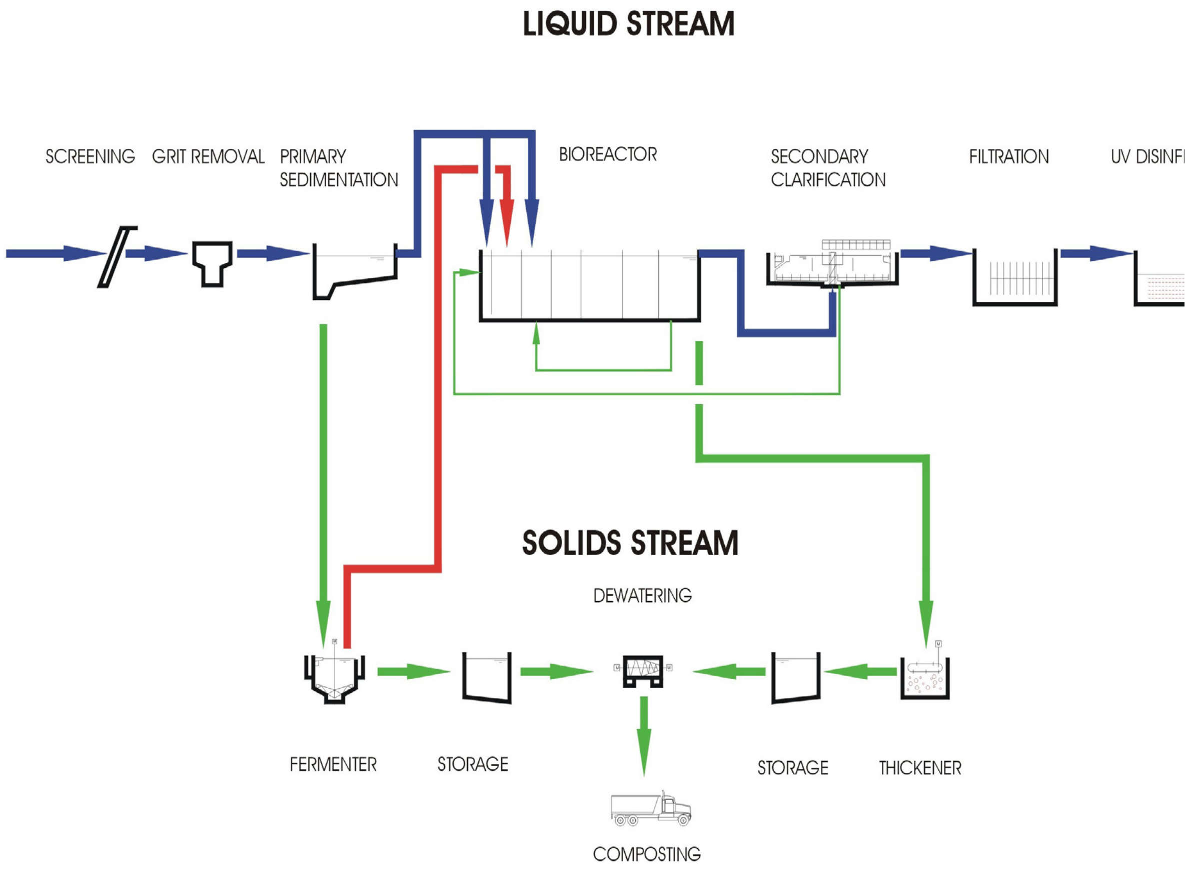

Figure 3. Kelowna WWTP 2005 configuration.

The Westbank process configuration employs a step-feed strategy for distributing primary effluent and fermenter supernatant (volatile fatty acids [VFA]-enriched) to the specific areas in the process where they are required. The logic described in the sections below was applied.

Return Sludge and Pre-anoxic Zone

The Kelowna secondary clarifier design included sidewall depths of 4 meters (m) or greater. Additionally, the original secondary clarifiers in Kelowna were designed with side-outlet stilling wells to reduce turbulence under the center inlet well. Floor sloping enabled sludge and helical scrapers to convey sludge to the center of the clarifier for collection and return to the bioreactor.

Typical return activated sludge (RAS) rates of 75 percent of the influent flow (Q) maintained sludge blankets of 0.5 to 0.75 m, which, when concentrated to three times the mixed liquor suspended solids (MLSS) concentrations, demonstrated significant denitrification potential.

Nitrate reductions in the RAS blanket of up to 6 mg/L have not caused rising sludge concerns; thus, the Kelowna secondary clarifiers have been operated since 1982 as anoxic denitrification zones and included in the overall process strategy.

With control of nitrates in the return sludge stream within the clarifier, there is minimal potential for nitrate return to the anaerobic zone. As an added protection, the original five-stage design included a small pre-anoxic zone for denitrification of any residual RAS nitrates before entering the anaerobic zone.

Given the limited potential for nitrate recycle in the return sludge, the amount of primary effluent required for RAS denitrification is greatly reduced. Plant personnel therefore developed plans to step-feed primary effluent to both the anaerobic zone (to stimulate phosphorus release) and the anoxic zones (to stimulate denitrification).

As a result of step-feeding the primary effluent to the main anoxic zone, the suspended solids concentration increases significantly in the pre-anoxic and anaerobic zones. With 50 percent primary effluent diversion, the suspended solids concentration is approximately 50 percent higher than MLSS concentrations in the aerobic zones.

With only a small amount of primary effluent added to the RAS entering the pre-anoxic zone, a very high denitrification rate ensures that no nitrate breaks through to the anaerobic zone.

The sizing of the pre-anoxic zone in Kelowna is less than 1 percent of bioreactor volume.

Anaerobic Zone

It has been well documented that the anaerobic zone requires consistent and sufficient VFA loadings to stimulate phosphorus release and uptake. The amount of VFA required has been documented as 4–8 kg VFA/kg soluble phosphorus removed.

At the Kelowna facility, a primary sludge fermenter was included in the original Bardenpho design, and it had a proven track record of consistent VFA production in the range required for good phosphorus removal. Therefore, the VFA-rich fermenter supernatant is discharged directly to the anaerobic zone, ensuring a steady feed of VFA to the phosphorus accumulation organisms (PAO).

It was established that with the side-stream VFA addition, the process performed better when the HRT of the anaerobic zones was reduced from 3 hours to 1 hour. This might have been the result of reduction of *secondary release of phosphorus* in the larger anaerobic cells.

With a Westbank configuration, the primary effluent step-feed to the anoxic zone is adjusted to complete two tasks:

- Primary effluent containing some VFA is added to the anaerobic zone along with the supernatant from the side-stream, primary-sludge fermenter. The combination of the two meets the total VFA requirements of the process.
- Primary effluent is step-fed to the anoxic zone to complete denitrification.

Under normal operating conditions, a portion of the primary effluent (approximately 50 percent) is required in the anoxic zone to complete denitrification, and the remainder is fed through the pre-anoxic zone to the anaerobic zone.

Anoxic Zone

The main anoxic zone requires a variable chemical oxygen demand (COD) load to control the denitrification process. Therefore, a portion of the primary effluent is pumped directly to the anoxic zone to stimulate denitrification. Using this technique, denitrification rates in the anoxic zone are greatly increased, the anoxic zones are reduced to 16–21 percent of bioreactor volume, and the amount of primary effluent step-feed to the anoxic zone is controlled.

Control of the denitrification rate can be achieved by monitoring the oxidation-reduction potential (ORP) at the end of the anoxic zone 24 hours a day. This information can be fed into the computer system and sufficient primary effluent diverted to the anoxic zone to meet the nitrate load from the nitrified internal recycle flow.

Aerobic Zone

The remaining volume (up to 75 percent) of the bioreactor is allocated for nitrification. This zone is sized on the basis of the nitrifier growth rate of the activated sludge during the coldest anticipated wastewater temperatures, and it controls the solids retention time (SRT) in the bioreactor.

One advantage of a step-fed configuration is the decrease in anaerobic and anoxic zone HRT—approximately 25 percent of the bioreactor. The reduced un-aerated fraction results in reducing the un-aerated decay rates for nitrifying bacteria. With shorter time spent under anoxic conditions, the net nitrifier growth rate increases. This is one reason for the reduced SRT normally used by plant operators in the Westbank configuration.

Table 3 provides a 2005 monthly summary of bioreactor operating parameters for HRT, SRT, temperature, MLSS, and percentage of bioreactor volume in service. Throughout 2005, one of the small modules was not required. In addition, the highest monthly MLSS was 2,803 mg/L, or approximately 80 percent of the design MLSS. It could be expected that an additional 20 percent load could be treated using the three operational bioreactors.

Table 3. Bioreactor operating parameters

Month	HRT (hr)	SRT (days)	Temp (°C)	MLSS (mg/L)	Bioreactor in service
Jan 2005	11.1	8.9	13.1	2,562	84%
Feb 2005	11.1	8.8	13.0	2,761	84%
Mar 2005	11.4	8.2	13.8	2,803	84%
Apr 2005	11.6	8.1	15.6	2,486	84%
May 2005	11.3	8.0	18.1	2,238	84%
Jun 2005	10.9	6.7	19.4	2,414	84%
Jul 2005	10.9	6.0	21.1	2,301	84%
Aug 2005	10.8	5.8	22.0	1,992	84%
Sept 2005	10.9	6.0	20.9	1,901	84%
Oct 2005	11.1	7.0	19.3	2,142	84%
Nov 2005	11.5	7.4	16.8	2,451	84%
Dec 2005	11.5	7.5	14.3	2,899	84%

Internal Nitrified Recycle Rates

Depending on the desired effluent nitrate concentration, the aerobic/anoxic configuration commonly uses four to six times the Q for internal recycle flows. With controlled primary effluent diversion to the anoxic zone, effluent nitrate concentrations in the 3.0 to 4.5 mg/L range can consistently be achieved.

The dissolved oxygen (DO) concentration in the aerobic zone can be reduced to between 1.0 to 2.0 mg/L with little impact on nitrifier growth rate, which is maximized at DO concentrations of 2.0 mg/L.

Three important advantages of reduced DO have assisted Kelowna operations:

- Reduced recycle of DO to the anoxic zone requires less primary effluent to initiate and complete denitrification.
- Reduced DO concentrations have reduced the endogenous release of nutrients.
- Reduced DO has reduced the proliferation of foam-producing organisms.

Supplemental Alum and Lime Addition

The Kelowna facility is equipped with a supplemental alum dosing system that is automated with an online analyzer. This system has been provided to help the biological phosphorus removal system achieve an annual average TP of 0.25 mg/L. The alum can be used if equipment maintenance or process issues disrupt effective phosphorus removals. As shown in Table 4, alum additions in 2005 were limited to 5 days.

The 1994 expansion included a lime system for controlling dissolved phosphorus in the centrifuge centrate return stream. The option of adding lime was terminated in March 2005 because of the strong bio-phosphorus removal performance in the bioreactor.

Table 4. Supplemental alum usage

2005	Alum (lb/d)
6/29/2005	500
6/30/2005	500
12/20/2005	150
12/21/2005	150
12/21/2005	200

Metals and Other Cations in Activated Sludge

Under normal operating conditions, the heavy-metal load to the Kelowna sewer system is typical of domestic sewage only. On rare occasions, however, discharges have disrupted both nitrogen and phosphorus removal. Throughout 2005, there were no such occasions, and

Table 5 shows typical metal concentrations found in the BNR sludge. With these concentrations of heavy metals, it could be expected that the nitrifier growth rate would be normal.

Table 5. Metals and other cations in activated sludge

Metal/Cation	Unit	Value	Metal/Cation	Unit	Value
Aluminum	µg/g	6,914	Manganese	µg/g	96.5
Antimony	µg/g	1.7	Mercury	µg/g	0.90
Arsenic	µg/g	1.9	Molybdenum	µg/g	6.88
Barium	µg/g	236	Nickel	µg/g	16.47
Beryllium	µg/g	0.11	Phosphorus	%	3.9
Bismuth	µg/g	12.27	Potassium	%	1.54
Cadmium	µg/g	1.40	Selenium	µg/g	4.34
Calcium	%	1.16	Silver	µg/g	11.07
Chromium	µg/g	17.77	Sodium	µg/g	2,446
Cobalt	µg/g	3.41	Strontium	µg/g	122.8
Copper	µg/g	768	Thallium	µg/g	0.309
Iron	µg/g	4,085	Tin	µg/g	3.78
Lead	µg/g	16.85	Vanadium	µg/g	7.18
Lithium	µg/g	2.37	Zinc	µg/g	288
Magnesium	%	1.08	Zirconium	µg/g	29.7

VFA Sources—Fermenter, Influent Sewage, Centrifuge

The primary sludge fermenter returns the overflow (supernatant) directly to the anaerobic zone of the bioreactor. Table 6 identifies the flows and concentrations of various parameters. As the data show, a significant amount of VFA is produced in the fermenter supernatant stream.

Table 6. Fermenter supernatant return to anaerobic zone

Month	Flow (mL/d)	SRT (d)	Solids (%)	Ammonia (mg/L)	Soluble phosphorus (mg/L)	Soluble COD (mg/L)	Suspended solids (mg/L)	Total VFA (mg/L)	VFA (kg/d)
Jan 2005	1.56	5.3	6.9%	18.7	5.62	358	142	116	181
Feb 2005	1.56	5.5	6.5%	18.1	6.39	407	170	131	204
Mar 2005	1.56	5.4	6.6%	18.6	7.06	427	169	140	218
Apr 2005	1.55	5.6	5.6%	19.5	7.68	538	180	196	305
May 2005	1.55	4.7	6.4%	15.8	7.61	632	166	225	351
Jun 2005	1.55	3.2	5.7%	15.7	6.95	583	190	236	368
Jul 2005	1.55	3.4	6.2%	14.4	7.85	640	212	254	393
Aug 2005	1.55	2.7	6.7%	16.4	7.16	611	208	242	375
Sept 2005	1.55	2.7	6.5%	16.5	6.82	575	227	229	355
Oct 2005	1.55	3.2	5.8%	18.6	7.43	582	232	222	344
Nov 2005	1.55	4.6	5.5%	19.3	7.92	603	198	216	334
Dec 2005	1.55	5.8	5.6%	20.5	7.85	614	260	227	351

Samples of the fermenter supernatant are sent off-site monthly for analysis in a gas chromatography (GC) analyzer to determine the concentration of various fractions of VFA. Table 7 lists the various fractional concentrations. The most desirable fraction for favoring the growth of PAOs is a combination of acetic and propionic acids stimulating phosphorus release/uptake. As the data show, these two acids are the most prevalent form of VFA in the fermenter supernatant.

Table 7. Fermenter VFA fractions

Month	Acetic (mg/L)	Propionic (mg/L)	Isobutyric (mg/L)	Butyric (mg/L)	Isovaleric (mg/L)	Valeric (mg/L)
Jan 2005	55.5	37.0	2.4	9.9	1.9	2.4
Feb 2005	65	26.1	2.2	9.7	2.1	2.6
Mar 2005	109	26.2	1.9	9	3.1	3.3
Apr 2005	154	57.8	1	26.8	1	9.9
May 2005	137	123	1.7	26.5	1.9	12.5
Jun 2005	121	64.5	2.6	21.2	1.1	6.3
Jul 2005	178	155	1.7	32	2	18.3
Aug 2005	209	105	3.8	24.9	3.4	9.3
Sept 2005	124	104	4.8	16.8	3.7	7.4
Oct 2005	165	105	1.9	2.7	1.7	9
Nov 2005	97	122	1	27.3	1	10
Dec 2005	122	130	3	33.3	2.7	14.6

VFAs are also found in the influent sewage and centrifuge centrate. Only a limited amount of sampling has been performed on these two sources. Table 8 lists the available data on influent, primary effluent, and centrifuge centrate VFA concentrations.

Given the limited number of samples, an estimate of the sources of VFA that feed the Kelowna anaerobic zone is as follows:

Primary sludge fermenter	Average 315 kg/d
50 percent of primary effluent	Average 252 kg/d

Table 8. Other VFA sources

Date	Centrifuge centrate (mg/L)	Primary effluent (mg/L)	Influent sewage (mg/L)	Flow rate	VFA (kg/d)
May 8, 2007	401			est. 130 m^3/d	52
May 10, 2007	285			est. 130 m^3/d	37
May 15, 2007	415			est. 130 m^3/d	54
June 8, 2006	281			est. 130 m^3/d	37
June 15, 2006	215			est. 130 m^3/d	28
June 22, 2006	281			est. 130 m^3/d	37
May 8, 2007		15		est. 36 ML/d	540
May 10, 2007		20		est. 36 ML/d	720
May 8, 2007			8	est. 32 ML/d	256

Centrifuge

The primary fermented and thickened waste activated sludge are combined at the centrifuge for dewatering and off-site composting. The key operating parameters for the centrifuge are included in Table 9. The first four months of 2005 included lime addition to the centrate to a level that saw the pH rise above 9.0. This effectively precipitated most of the soluble phosphorus to low levels. In May 2005 the operations staff stopped adding lime to the centrate because the bio-phosphorus removal efficiencies in the bioreactor were such that the return phosphorus load was effectively removed biologically and the assistance provided by lime addition was not required.

Table 9. Centrifuge centrate return to plant influent

Flow	Flow (m^3/d)	Ammonia (mg/L)	TP (mg/L)	Soluble P (mg/L)	TKN (mg/L)	Soluble COD (mg/L)	TSS (mg/L)
Jan 2005	133.7	14.3	118	11	43	318	1,105
Feb 2005	113.6	15.3	70	23	46	376	1,115
Mar 2005	124.5	15.3	160	55	68	452	861
Apr 2005	117.3	17.5	91	47	54	522	1,045
May 2005	118.7	16.7	225	173	63	827	270
Jun 2005	128.8	23.1	235	159	54	667	320
Jul 2005	143.6	28.5	165	161	60	783	1,001
Aug 2005	124.5	22.4	200	164	55	726	520
Sept 2005	118.7	23.5	235	148	61	599	1,135
Oct 2005	128.5	17.6	200	96	95	632	1,084
Nov 2005	123.8	20.9	173	118	66	779	854
Dec 2005	135.3	21.4	170	84	95	593	939

Performance Data for Nitrogen Removal

Overall plant influent and final filtered effluent average results for the 2005 calendar year are shown in Table 10. The operators at the Kelowna facility have found that to maximize biological phosphorus removal, the SRT needs to be *just enough to complete nitrification*.

If a small amount of ammonia remains in the effluent (0.2–0.5 mg/L), biological phosphorus removal appears to work at top efficiency. Table 10 shows the monthly averages in 2005, achieved as a result of this strategy. Tables 11 and 12 show the nitrogen concentrations at various stages in the process.

Table 10. Nitrogen removal

Month	Influent flow (ML/d)	Influent TKN (mg/L)	Effluent TN (mg/L)	Nitrogen removal (%)	Effluent nitrates (mg/L)	Effluent ammonia (mg/L)
Jan 2005	33.2	30.6	4.64	84.8	2.10	0.85
Feb 2005	32.1	30.5	4.90	83.9	1.93	1.01
Mar 2005	31.5	27.0	4.40	83.7	2.20	0.51
Apr 2005	30.8	32.5	4.49	86.1	2.65	0.48
May 2005	32.3	24.0	4.12	81.3	2.21	0.51
Jun 2005	32.8	24.7	3.21	87.0	1.99	0.07
Jul 2005	33.0	27.0	3.53	86.9	2.08	0.44
Aug 2005	33.5	33.0	4.39	86.6	2.53	0.40
Sept 2005	33.4	27.5	4.45	83.8	2.80	0.52
Oct 2005	32.3	27.8	4.89	82.4	2.67	0.50
Nov 2005	31.2	30.7	4.66	84.8	2.45	0.67
Dec 2005	31.9	31.1	4.78	84.6	2.18	0.96

Table 11. Nitrate profile—annual average of grab samples taken at 8:00 a.m. (mg/L)

Anaerobic zone	End Anoxic	25% aerobic	50% Aerobic	End aerobic	Secondary clarifier	Return sludge	Filter effluent
0.02	0.2	1.1	1.9	2.8	2.5	0.13	2.6

Table 12. Ammonia profile—annual average of grab samples taken at 8:00 a.m. (mg/L)

Primary effluent	Anaerobic Zone	End anoxic	25% aerobic	50% aerobic	End aerobic	Secondary clarifier	Return sludge	Filter effluent
19.44	9.6	3.19	2.12	1.31	0.05	0.29	0.26	0.23

Figures 4 and 5 show monthly frequency curves for effluent TN and ammonia.

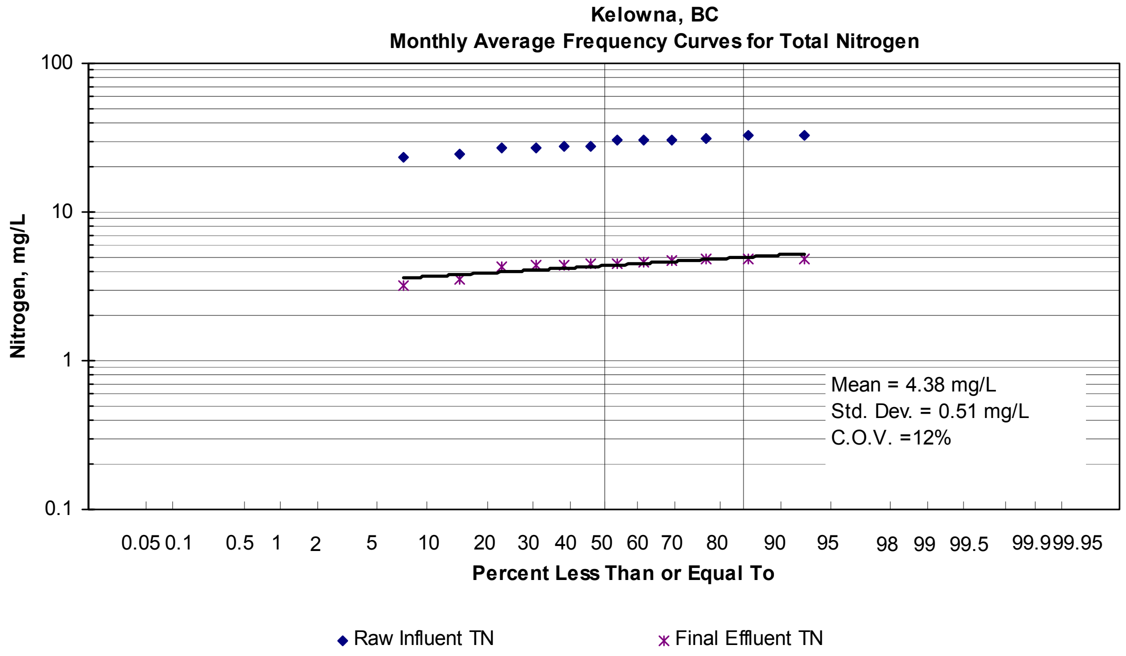

Figure 4. Monthly frequency curves for effluent TN.

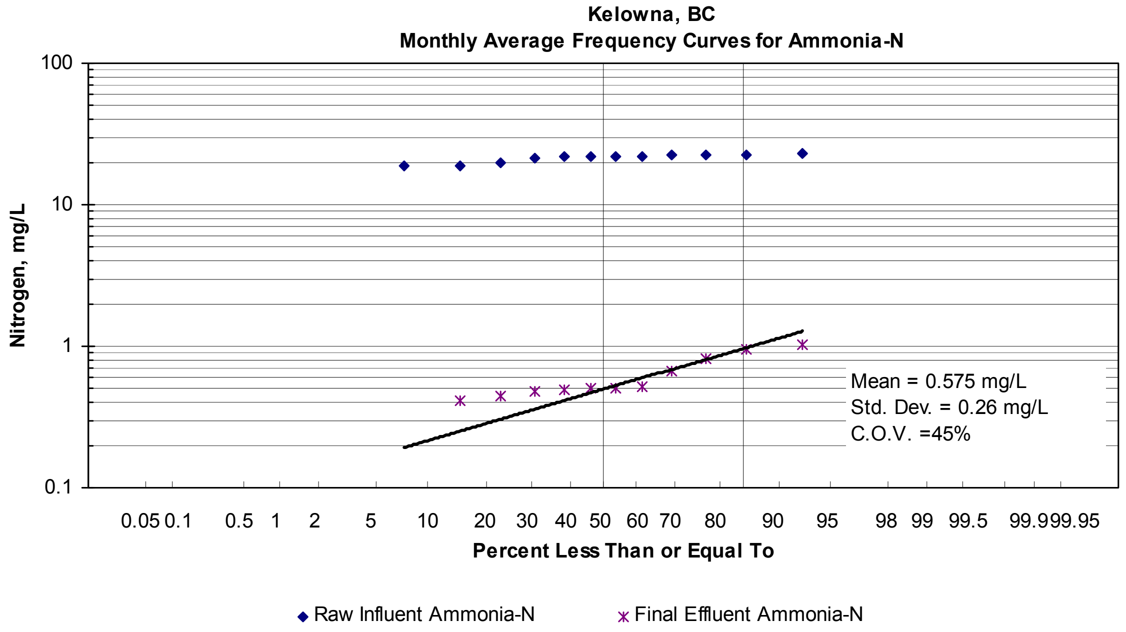

Figure 5. Monthly frequency curves effluent ammonia.

Performance Data for Phosphorus Removal

Overall plant influent and final filtered effluent average results for the 2005 calendar year are shown in Table 14. As the data show, biological removal of soluble phosphorus is operating at near maximum capability.

Table 14. Phosphorus removal

Date	Influent flow (ML/d)	Influent TP (mg/L)	Effluent TP (mg/L)	Phosphorus removal (%)	Effluent soluble P (mg/L)
Jan 2005	33.2	5.9	0.13	97.8%	0.04
Feb 2005	32.1	5.95	0.16	97.3%	0.04
Mar 2005	31.5	5.5	0.16	97.1%	0.04
Apr 2005	30.8	7.35	0.13	98.2%	0.04
May 2005	32.3	5.67	0.19	96.6%	0.05
Jun 2005	32.8	5.4	0.11	97.9%	0.03
Jul 2005	33.0	6.05	0.12	98.0%	0.03
Aug 2005	33.5	6.1	0.10	98.3%	0.03
Sept 2005	33.4	6.3	0.10	98.4%	0.02
Oct 2005	32.3	6.35	0.12	98.1%	0.02
Nov 2005	31.2	5.03	0.13	97.4%	0.02
Dec 2005	31.9	6.15	0.21	96.5%	0.06

Table 15 shows the soluble phosphorus concentrations at various stages in the process.

Table 15. Ortho-phosphorus profile—annual average of grab samples taken at 8:00 a.m. (mg/L)

Primary Effluent	Anaerobic zone	End anoxic	25% aerobic cell	50% aerobic	End aerobic	Secondary clarifier	Return sludge	Filter effluent
4.26	14.9	2.54	0.18	0.01	0.01	0.02	1.91	0.03

The soluble phosphorus load to the aerobic zone is quite low because of the moderate release of phosphorus in the anaerobic zone and the significant phosphorus uptake in the anoxic zone for most of the year.

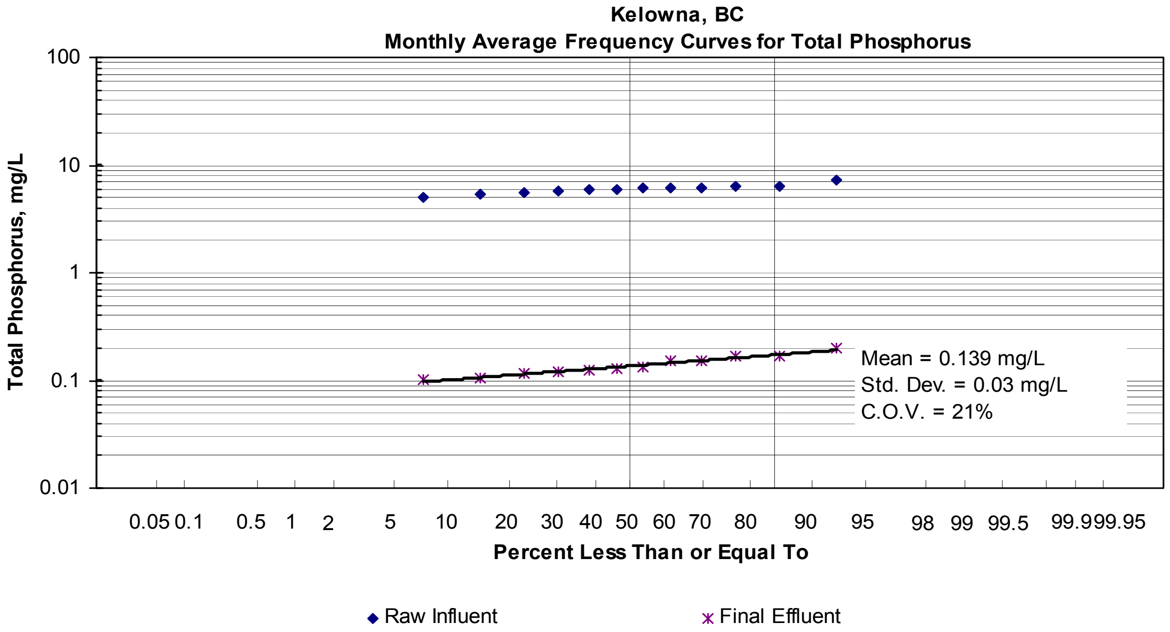

Figure 6. Monthly average frequency curves for TP.

Reliability Factors

The Kelowna plant has achieved a high degree of reliability in the biological removal of nitrogen and phosphorus in a cold climate. The mean effluent concentrations were 0.14 mg/L in TP with a low coefficient of variation (COV) of 21 percent and 4.38 mg/L in TN with a low COV of 12 percent.

The key operating principles applied at the Kelowna site include the following:

- Anaerobic zone sizing was reduced from 3 hours to 1 hour for optimal operation when a primary sludge fermenter was used to produce a constant, side-stream VFA source.
- Secondary clarifiers with a bottom-central draw-off are used to significantly reduce nitrates in the return sludge.
- The secondary clarifier RAS rate is adjusted to remove nitrates and prevent excessive phosphorus release.
- A small pre-anoxic zone for final denitrification of RAS before entering the anaerobic zone prevents excessive phosphorus release before the anaerobic zone.
- When a portion of the primary effluent was introduced directly to the anoxic zone, rapid denitrification occurred and anoxic zone sizing could be reduced.

- Simultaneous nitrification/denitrification occurred when submerged turbine aerators were used, thereby improving the overall nitrate removal.
- DO in the range of 1.0 to 2.0 mg/L produced the best combined TN and TP removals.
- Sufficient SRT is maintained to just achieve full nitrification. A small amount of ammonia in the effluent is acceptable.
- Online effluent monitoring of nutrients provides valuable information to the plant operators.

  If there is a soluble phosphorus breakthrough to the effluent, the online effluent analyzer that collects and analyzes samples every 15 minutes for ammonia, nitrates, and ortho-phosphorus provides a signal to the process computer, which can automatically turn the supplemental alum-dosing upstream of the secondary clarifiers on or off.

- Flow and load equalization volume equivalent to 7.5 percent of daily flow helps to stabilize the nutrient removal processes.
- With a 6Q recycle, the fourth and fifth stages in the five-stage Bardenpho mode were not required to meet TN and TP permit requirements.
- Computer control systems monitor, operate, and alarm all equipment on-site. This provides 24-hour-a-day, consistent process control.
- The anoxic zone is removing significant amounts of dissolved phosphorus. This appears to be stimulated by the addition of primary effluent and the higher denitrification rates.
- Recycle loads from dewatering were minimized by maintaining separate processes for secondary sludge and primary sludge. No sludge digestion was practiced in Kelowna. The total recycle loads from dewatering were only 13 percent in TP and 0.1 percent in TN.
- Wet-weather flows were managed under the normal mode of operation, using the equalization basin. The sewer system was separated, and the seasonal variation in flow was not very high. The maximum month flow was 10 percent higher than the average flow. The total basin equalization capacity was 7.5 percent of the design average flow.

All these operating principles have been put into effect because of the flexibility of process layout, the built-in swing zones, and the leadership of the plant personnel in research and process optimization.

Costs

Treatment Plant Expansions

This section provides a design summary of the Kelowna facility expansions, from the 1980 expansion and Bardenpho upgrade (from conventional, high-rate activated sludge) through two additional upgrades to the Westbank process—each with higher loadings than the original Bardenpho bioreactor.

Expansion of 1969 Kelowna WWTP

The Kelowna WWTP was converted in 1980 from secondary treatment to nutrient removal. The following facilities from the previous 1969 expansion were incorporated into the design:

- Two influent comminutors
- Two grit channels
- Raw sewage lift station
- Three primary clarifiers
- Short HRT activated-sludge process (converted to flow equalization)
- Two secondary clarifiers (converted to sludge fermenters in Phase 2)
- Sludge thickener

1980 Five-stage Bardenpho

The 1980 Bardenpho five-stage design made the Kelowna WWTP the first full-scale facility designed for nutrient removal in North America. The unique and highly flexible bioreactor had two trains, each with 22 cells for anaerobic, anoxic, and aerobic service. Of the 22 cells, 17 were *swing zones* with either anoxic or aerobic configurations. This design enabled complete flexibility in operating the nitrifying and denitrifying components of the process.

The original design was commissioned with a high priority on reliability. Consequently, a very conservative HRT/SRT was used to ensure complete nitrification and denitrification to facilitate a TN below 6.0 mg/L. Through extended optimization, it became clear that the long HRT/SRT was not necessary to achieve the required effluent nitrogen standards.

The preexisting sludge thickener was put into service for primary sludge only with supernatant returning to the influent works. Thus it provided sufficient *rapidly degradable COD* to stimulate phosphorus removal and denitrification. Through extended optimization, it became clear that the on-site thickener (later called a *fermenter*) was producing sufficient VFA to reduce the anaerobic zone from three cells to a single cell.

The capital cost for the 1980 conversion to the Bardenpho configuration was 12.5 million Canadian dollars (CDN$).

Westbank Process Configuration

On the basis of the full-scale operation of the five-stage process, a more compact process was developed and initially tested at Kelowna. Then a full-scale version was designed and constructed at the Westbank WWTP site across the lake from Kelowna.

The Westbank configuration uses a step-feed primary effluent strategy to split the primary effluent (COD) for denitrification in the anoxic zones. It also ensures anaerobic conditions for phosphorus release in the anaerobic zone.

Using a primary sludge fermenter with direct discharge to the anaerobic zone provides a consistent VFA source, and primary effluent is added to the anaerobic zone only if additional VFA load to the anaerobic zone is required.

The *high-rate* Westbank process was implemented in two phases. The first phase involved breaking up the five-stage process into two intermediate-sized bioreactors and two smaller bioreactors. The second phase added more capacity upstream and downstream to the original bioreactor.

The objective of the second-phase expansion was to fully develop the capacity of the original bioreactor with the new high-rate process. The plant was again re-rated upward to an average dry weather flow of 10.6 million gallons per day (MGD) (40 ML/d).

The principal change to the process involved a controlled diversion of primary effluent to enhance the denitrification rate in the main anoxic zone. The addition of primary effluent directly to the anoxic zone allowed smaller anoxic zones and facilitated adjustment to the denitrification rate. Combined with the smaller anaerobic zone previously developed in the 1980s, the aerobic fraction of the process was increased from 55 percent to 71 percent.

The capital cost of the 1992 Phase 2 conversion was approximately CDN$6.2 million. The capital cost of the 1994 Phase 3 conversion was approximately CDN$20.75 million.

Canadian–U.S. Dollar Exchange

To calculate the capital and operation and maintenance (O&M) costs in U.S. dollars, Canadian-to-U.S. dollar exchange rate values were required. Table 16 presents the average Canadian–to-U.S. dollar exchange rates in the 3 years that capital improvements were made, along with the current exchange rate for calculating O&M costs (Oanda Corporation 2007).

Table 16. Average Canadian to U.S. dollar exchange rate value

Year	1 Canadian $ = × U.S. $
1980	0.86
1992	0.83
1994	0.73
2007	0.94

Table 17 presents the assumed split of the capital cost among phosphorus removal, nitrogen removal, and other, which is BOD removal. It was assumed that 12 percent of the upgrades could be attributed to phosphorus removal, while 48 percent of the upgrades could be attributed to nitrogen removal. The balance of the upgrades could be attributed to BOD removal or other activities required by permit (e.g., filters for suspended solids). This meant that the capital expenditure in 2007 dollars that could be attributed to phosphorus removal was US$6.8 million. The annualized capital charge (20 years at 6 percent) was US$595,000 for phosphorus removal.

Table 17. Split of capital cost between phosphorus, nitrogen, and other

Capital year	CDN$	US$	US$ present worth	% other	%P	%N	Phosphorus	Nitrogen
1980	$12,500,000	$10,750,000	$26,375,193	40%	12%	48%	$3,165,023	$12,660,093
1992	$6,200,000	$5,146,000	$8,198,502	40%	12%	48%	$983,820	$3,935,281
1994	$20,750,000	$15,147,500	$22,245,090	40%	12%	48%	$2,669,411	$10,677,643
Totals	$39,450,000	$31,043,500	$56,818,785				$6,818,254	$27,273,017

The capital expenditure in 2007 dollars that could be attributed to nitrogen removal was US$27.2 million. The annualized capital charge (20 years at 6 percent) was US$2.38 million. for nitrogen removal. This same expenditure could be attributed to ammonia nitrogen removal.

The total capital attributed to BNR in 2007 dollars was US$34 million. For the 10.6 MGD (40 ML/day) facility, this means the capital expenditure per gallon of BNR treatment capacity was US$3.25.

Operation and Maintenance Costs

The plant uses both biological phosphorus and nitrogen removal, with minimal use of alum and no use of supplemental carbon sources. This means that costs for nutrient removal are essentially all electrical. A summary of the electrical calculations is provided in Attachment 2. The total electrical usage for phosphorus removal was 884,000 kilowatt-hours per year (kWh/yr). When the average electrical rate of US$0.047/kWh was applied, the cost for phosphorus removal was US$41,500 for the year. The total electrical usage for nitrogen removal was 4,100,000 kWh/yr, or US$193,000.

Unit Costs for Nitrogen and Phosphorus Removal

During the 1-year case study period, the plant removed 150,000 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal is US$0.27 and the unit capital cost is US$3.97/lb of phosphorus removed.

During the same period, the plant removed 781,000 lb of TN. With the results above, the unit O&M cost for TN removal is US$0.14 and the unit capital cost is US$3.05/lb of ammonia removed.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal is US$4.24/lb phosphorus removed and the life-cycle cost for nitrogen removal is US$3.19/lb TN removed.

Assessment of magnitude of costs: The costs are shown to be on the high side in capital cost and very low in O&M costs. This reflects the innovative technologies used at the plant, which resulted in increasing the treatment capacity while still using the existing facilities.

Summary

The Kelowna, British Columbia, plant's retrofit of the original five-stage Bardenpho process into the three-stage Westbank process has provided excellent reliability in both nitrogen and phosphorus removal, especially for this cold-weather region. The phosphorus removal is achieved biologically to the mean concentration of 0.14 mg/L with a low COV of 21 percent. The nitrogen removal is achieved biologically to the mean concentration of 4.38 mg/L with an extremely low COV of 12 percent without using an external carbon source. The Kelowna plant is one of the best-performing BNR plants in North America. Many factors have contributed to this remarkable achievement. They include flexibility in design for bioreactors, adequate VFA production in separate fermenters, online monitoring and automatic controls, and the plant personnel developing optimal operating strategies.

Key factors include downsizing the anoxic zones; maintaining 2- to 3-foot-deep blankets in the secondary clarifier for added denitrification, thereby downsizing the pre-anoxic zone; simultaneous nitrification and denitrification; DO controls in the range of 1 to 2 mg/L in the aerobic zone; maintaining a short sludge age of about 10 days, a short HRT of about 11 hours, and sufficient internal recirculation for denitrification at 6Q; and a computer control system. Recycle loads from sludge handling were minimized by maintaining separate processes for secondary sludge and primary sludge. No sludge digestion was practiced, and thus the total recycle loads were 13 percent in TP.

The capital cost was moderately high at US$3.25 per gallon per day, and the O&M costs were extremely low at US$0.28/lb of phosphorus removed and $ US$0.29/lb of nitrogen removed. The capital cost reflects added costs for flexible flow patterns with multiple swing zones for both anoxic and aerobic zones, fermenters, and tertiary filters. The O&M costs are low because of efficient use of power and no chemical addition for either nitrogen or phosphorus removal. The life-cycle costs are low at US$3.19/lb of nitrogen and US$4.25/lb of phosphorus removed.

As a result of the continuous improvements, the Kelowna plant treats 70 percent more flow than the original plant did using the same bioreactor tanks.

Acknowledgments

The authors are grateful to Earth Tech Canada, Ltd., and Gerry Stevens, Earth Tech Canada, Ltd., for the bulk of the information contained in this case study. Mr. Stevens worked for Kelowna through implementing and optimizing the BNR process until 1990. He helped to obtain records and analyze and present technical data about the innovations achieved at the plant. Mr. Stevens' expertise in BNR technologies in general and his accomplishments in Kelowna, British Columbia, are recognized and appreciated.

The authors are also grateful to the city of Kelowna and members of the WWTP staff for participating in this study. They include Jim White, plant manager; Sheila Carey, lab supervisor; and Marj Van de Mortel, lab technician.

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Attachment 1: Design Basis

Design Flows and Loads	Units	1980 Bardenpho upgrade	Design value	
			Stage 1	
Flow Data		Phase 1	Phase 2	Phase 3
Sewered Population	per	56,000	64,000	95,000
Flow per Capita	L/c.d	400	400	400
Base Infiltration	ML/d	1.0	1.0	1.0
BCTWP Industrial Effluent	ML/d			1.0
Average Daily Flow	ML/d	22.5	27.5	40.0
Maximum Month Flow	ML/d	25.1	31.0	44.0
Maximum Daily Flow	ML/d	28.7	35.5	50.0
Peak Hourly Flow	ML/d	34.8	43.0	69.0
BOD, TSS, TKN, TP Loads				
BOD				
Average Daily Unit Load	kg/c.d	0.080	0.080	0.080
Allowance for BCTWP	kg/d			200
Average Daily Total	kg/d	4,480	5,120	7,800
Maximum Month Unit Load	kg/c.d	0.095	0.095	0.095
Maximum Month Total	kg/d	5,320	6,080	9,225
Maximum Week Unit Load	kg/c.d	0.105	0.105	0.105
Maximum Week Total	kg/d	5,880	6,720	10,175
TSS				
Average Daily Unit Load	kg/c.d	0.080	0.080	0.080
Allowance for BCTWP	kg/d			20
Average Daily Total	kg/d	4,480	5,120	7,620
Maximum Month Unit Load	kg/c.d	0.100	0.100	0.100
Maximum Month Total	kg/d	5,600	6,400	9,520
Maximum Week Unit Load	kg/c.d	0.120	0.120	0.120
Maximum Week Total	kg/d	6,720	7,680	11,420
TKN				
Average Daily Unit Load	kg/c.d	0.015	0.015	0.015
Allowance for BCTWP	kg/d			10
Average Daily Total	kg/d	840	960	1,435
Maximum Month Unit Load	kg/c.d	0.017	0.017	0.017
Maximum Month Total	kg/d	952	1,090	1,625
Maximum Week Unit Load	kg/c.d	0.019	0.019	0.019
Maximum Week Total	kg/d	1,064	1,215	1,815
TP				
Average Daily Unit Load	kg/c.d	0.003	0.003	0.003
Allowance for BCTWP	kg/d			5
Average Daily Total	kg/d	168	192	290
Maximum Month Unit Load	kg/c.d	0.003	0.003	0.003
Maximum Month Total	kg/d	168	192	290
Maximum Week Unit Load	kg/c.d	0.004	0.004	0.004
Maximum Week Total	kg/d	224	256	385
WASTEWATER TEMPS				
Summer	°C	20	20	20
Winter	°C	10	10	10

Process Design Data	Units	1980 Bardenpho upgrade	Design value	
			Stage 1–upgrade	
		Phase 1	Phase 2	Phase 3
Raw Sewage Pumping				
Station 1				
Number of Units		6	6	
Capacity	L/s	380	380	
Station 2				
Number of Units				3
Capacity	L/s			440
Comminutor				
Number of Units				
Mechanical		2		
Manual		--		
Capacity per unit	ML/d	16.0		
Bar Screen				
Number of Units				
Mechanical			1	1
Manual			1	1
Capacity per unit	ML/d		75.0	75.0
Grit Removal				
Number of Units		2	1	1
Capacity per Unit	ML/d	16.0	75.0	75.0
Primary Clarifiers				
Number of Units		3	4	6
Length	m	27.4	27.4	27.4
Width	m	6.1	6.1	6.1
SWD, 1-3	m	2.2	2.2	2.2
SWD, 4-6	m	2.5	2.5	2.5
SWD, 7-10	m			
Peak OFR, 1 out of service	$m^3/m^2 \cdot d$	62.7	94.1	91.1
Primary Flow Equalization				
Fraction of Average Flow	percent	8.4	6.9	7.50
Volumes				
NE Trunk	m^3	1,200	1,200	--
Existing Tanks	m^3	700	700	700
Future Primary Clarifiers	m^3		--	1,150
New Equalization Tanks	m^3		--	1,200
Primary Sludge Fermenters				
SRT, avg	d	7	5	5
Number of Units		1	1	2
Dimensions				
Diameter	m	17	15	15
SWD, 1-2	m	4.5	3.5	3.5
SWD, 3-4	m	--	--	--

Process Design Data	Units	1980 Bardenpho upgrade	Design value	
			Stage 1–upgrade	
		Phase 1	Phase 2	Phase 3
Bioreactors				
Basic Design Parameters				
SRT, Summer	d	15	12	10
SRT, Winter	d	20	15	12
Bioreactor				
Existing Modules 1 and 2				
No. of Anaerobic Cells		3		
Anaerobic Volume	m^3	1,365		
No. of Anoxic Cells		6–10		
Anoxic Volume	m^3	3,640		
No. of Aerobic Cells		9–13		
Aerobic Volume	m^3	5,005		
No. of Anaerobic Stirrers		3		
Anaerobic Stirrer hp		5		
No. of Swing Zone Mixers		19		
Swing Zone Mixers hp		7.5/15		
Bioreactor				
Modified Modules 1 and 4				
No. of Anaerobic Cells			1	1
No. of Anaerobic Stirrers			1	1
Anaerobic Stirrer hp			5	5
Anaerobic Volume	m^3		225	225
No. of Anoxic Cells			2	2
No. of Anoxic Stirrers			2	2
Anaerobic Stirrer hp			5	5
Anoxic Volume	m^3		680	680
No. of Aerobic Cells			4	4
Aerobic Volume	m^3		1,820	1,820
No. of Anaerobic Stirrers			1	1
Anaerobic Stirrer hp			2.5	2.5
No. of Aerobic Mixers			1	1
Aerobic Mixer hp			40	40
No. of Aerobic Mixers			1	1
Aerobic Mixer hp			30	30
No. of Swing Zone Mixers			3	3
Swing Zone Mixers hp			7.5/15	7.5/15

Process Design Data	Units	1980 Bardenpho Upgrade	Design Value	
			Stage 1–Upgrade	
		Phase 1	Phase 2	Phase 3
Modified Modules 2 and 3				
No. of Anaerobic Cells			1	1
Anaerobic Volume	m^3		455	455
No. of Anoxic Cells			3	3
Anoxic Volume	m^3		1,365	1,365
No. of Aerobic Cells			10	10
Aerobic Volume	m^3		4,550	4,550
No. of Anaerobic Stirrers			1	1
Anaerobic Stirrer hp			5	5
No. of Aerobic Mixers			2	2
Aerobic Mixer hp			40	40
No. of Aerobic Mixers				2
Aerobic Mixer hp				30
No. of Swing Zone Mixers			7	5
Swing Zone Mixers hp			7.5/15	7.5/15
Blowers				
No. of Blowers		4	4	4
Size	hp	100	100	250

Process Design Data	Units	1980 Bardenpho Upgrade	Design Value	
			Stage 1–Upgrade	
		Phase 1	Phase 2	Phase 3
Secondary Clarifiers				
Clarifiers				
Number		3	4	5
Dimensions				
Diameter	m	26	26	26
SWD	m	4.5	4.5	4.5
RAS Pumps				
Number		6	8	9
Capacity	L/s	80	80	80
Maximum RAS Flow	L/s	240	320	400
WAS Pumps				
Number		2	3	4
Capacity	L/s	8	12	12
Filtration				
Peak OFR, 1 unit out of service		290	290	290
Existing Units				
Number		4	4	4
Area per Unit	m^2	64	64	64
New Units				
Number				1
Area per Unit	m^2			96
Ultraviolet Disinfection				
Dosage	mWs/cm^2	chlorine	chlorine	48
Transmissivity	percent			65
Number of Lamps				1,152
Arrangement				
Number of Channels				2
Banks per Channel				3
Racks per Bank				24
Lamps per Rack				8
WAS Thickening				
Design Load, Peak	kgTSS/d			4,615
DAF Units				
Number		2	2	3
Area per Unit	m^2	18.9	18.9	18.9
Dewatering				
PS Flow, peak	m^3/d	none	none	90
WAS Flow, Peak	m^3/d	none	none	195
Centrifuges				
Number				2
Capacity	L/s	none	none	4.7

Attachment 2: Electrical Cost

Electrical cost

Anoxic/Anaerobic mixers

HP	Number	kW power draw	hours/ day	kWh draw/ day	kWh draw/ year	%BOD	%P	%N	for P draw	for N draw
Anaerobic mixer										
5	2	7.46	24	179.04	65,349.6	0	100	0	65,349.6	0
2.5	1	1.865	24	44.76	16,337.4	0	100	0	16,337.4	0
Fermenter rake mechanism drive										
5	1	3.73	24	89.52	32,674.8	0	100	0	32,674.8	0
Anoxic mixers										
5	8	29.84	24	716.16	261,398.4	0	0	100	0	261,398.4
2.5	2	3.73	24	89.52	32,674.8	0	0	100	0	32,674.8
Blowers										
250	1.25	233.125	24	5,595	2,042,175	45	10	45	204,217.5	918,978.75
Swing zone stirrers—19 available, can go either anoxic (7.5 hp) or aerobic (15 hp)										
7.5	9	50.355	24	1,208.52	441,109.8	0	0	100	0	441,109.8
15	10	111.9	24	2,685.6	980,244	45	10	45	98,024.4	441,109.8
Aerobic zone mixers										
40	5	149.2	24	3,580.8	1,306,992	45	10	45	130,699.2	588,146.4
30	5	111.9	24	2,685.6	980,244	45	10	45	98,024.4	441,109.8
15	11	123.09	24	2,954.16	1,078,268.4	45	10	45	107,826.84	485,220.78
Recirculation pump										
20	2	29.84	24	716.16	261,398.4	0	0	100	0	261,398.4
15	1	11.19	24	268.56	98,024.4	0	0	100	0	98,024.4
Filter pumps										
7.5	4	22.38	24	537.12	196,048.8	0	50	50	98,024.4	98,024.4
10	1	7.46	24	179.04	65,349.6	0	50	50	32,674.8	32,674.8

Marshall Street Water Reclamation Facility Clearwater, Florida

Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

The Marshall Street Water Reclamation Facility (WRF) in Clearwater, Florida, is designed for a capacity of 10 million gallons per day (MGD). This facility was selected as a case study because it has achieved low levels of nitrogen and phosphorus in the effluent using the five-stage Bardenpho process. The plant processed an average of 5.48 MGD during the evaluation period, October 2005 through September 2006. Some of the reclaimed water is sent for reuse (irrigation); the remainder is discharged under a permit via Stevenson's Creek to Clearwater Harbor. The WRF uses a five-stage Bardenpho process to remove both total nitrogen (TN) and total phosphorus (TP) to below 3 milligrams per liter (mg/L) and 1 mg/L on an annual average, respectively.

The relevant National Pollutant Discharge Elimination System (NPDES) permit limits for the facility are shown in Table 1.

Table 1. NPDES permitted discharge limits

Parameter	Annual average	Monthly average	Weekly average
BOD_5	5 mg/L	6.25 mg/L	7.5 mg/L
TSS	5 mg/L	6.25 mg/L	7.5 mg/L
TN	3 mg/L	3.75 mg/L	4.5 mg/L
TP	1 mg/L	1.25 mg/L	1.5 mg/L
Dichlorobromo-methane	24 µg/L	Report	--
Dibromochloro-methane	46 µg/L	Report	--

Notes:
µg/L = micrograms per liter
BOD_5 = biochemical oxygen demand
TSS = total suspended solids
TN = total nitrogen
TP = total phosphorus

Plant Process

Figures 1 and 2 present a plant layout and a process flow diagram for the Marshall Street WRF.

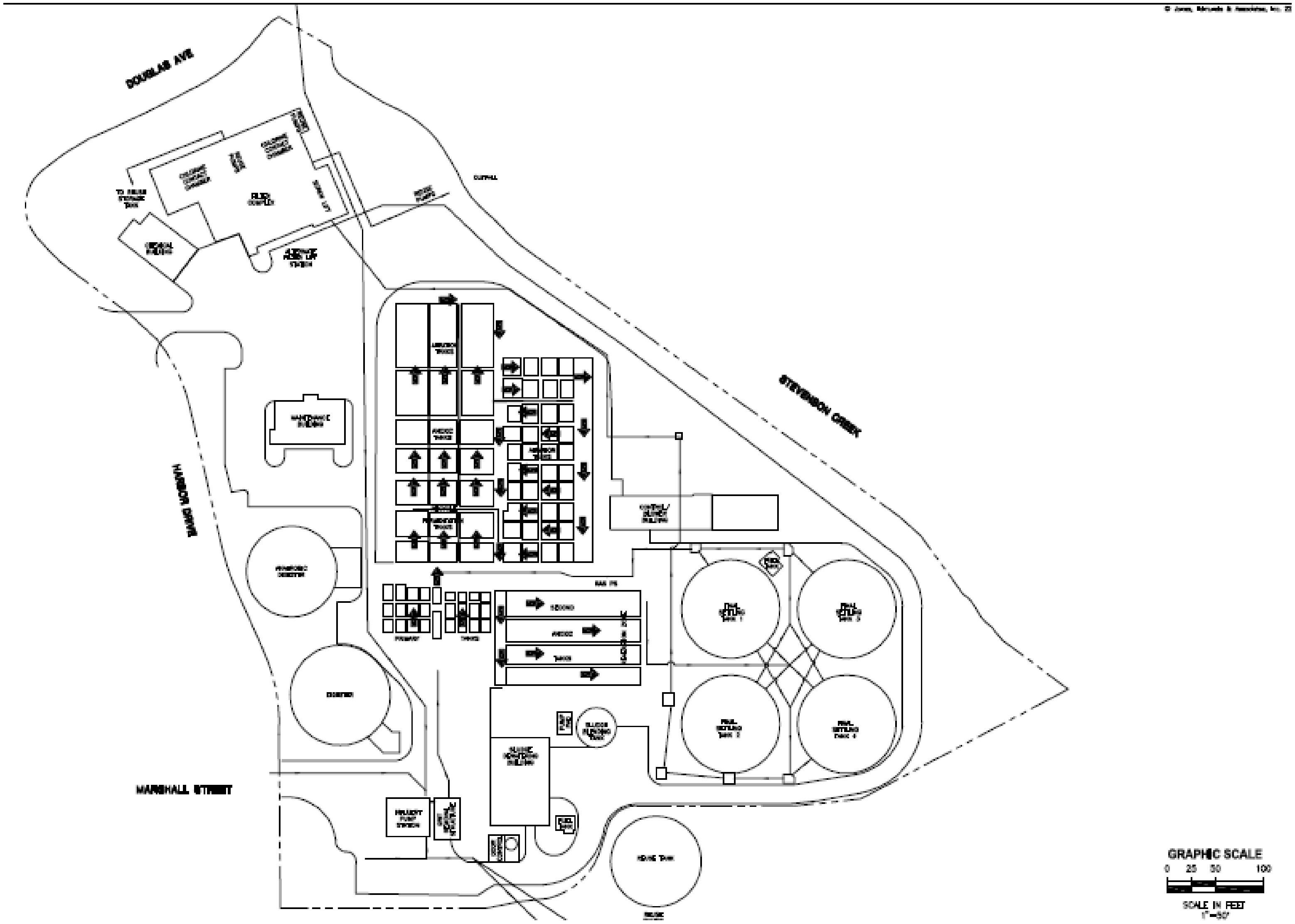

Figure 1. Marshall Street WRF layout.

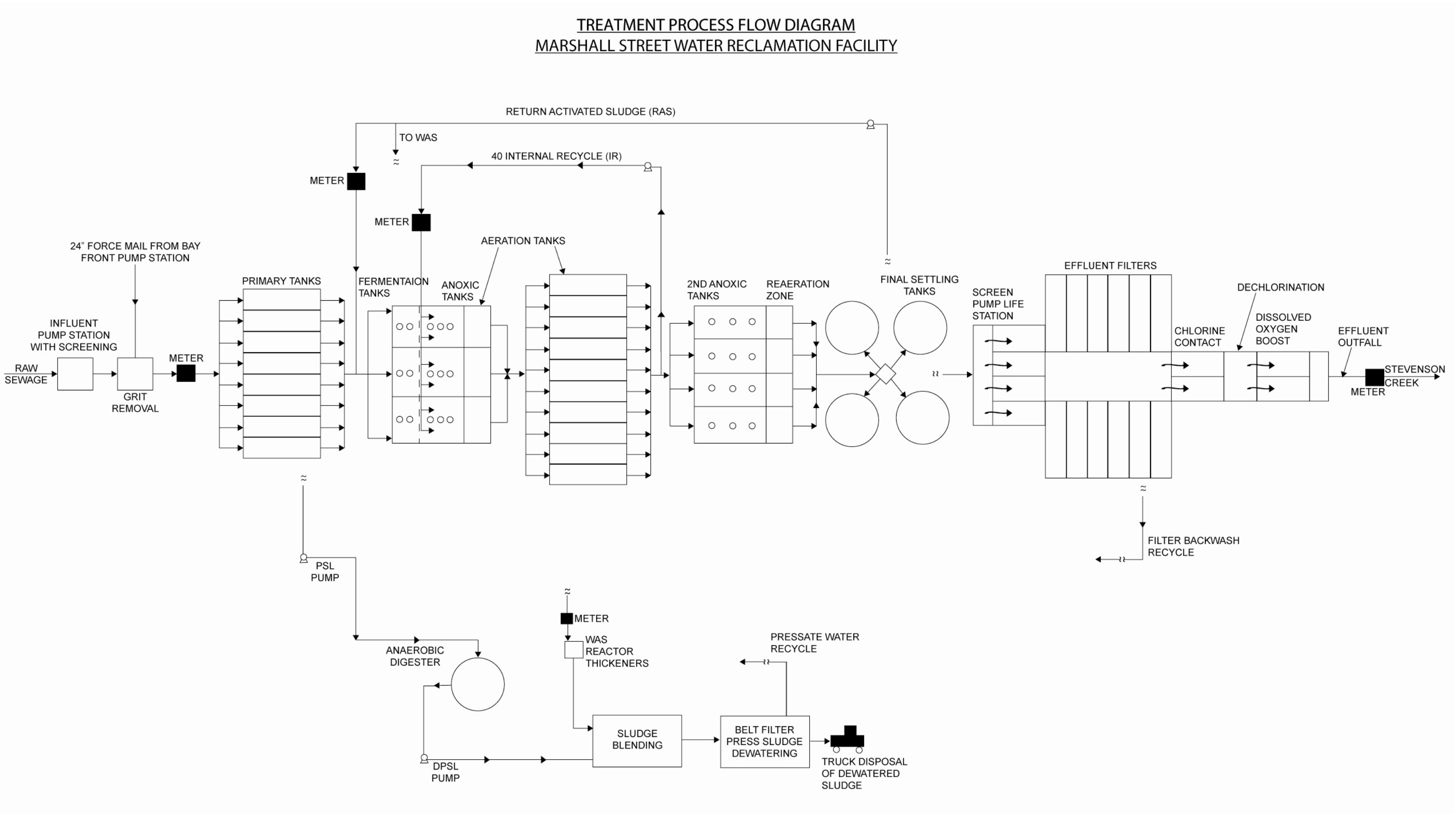

Figure 2. Marshall Street WRF process flow diagram.

The plant uses a five-stage Bardenpho biological nutrient removal (BNR) process. The liquid train consists of the following components: an on-site influent pumping station with three variable-rate, dry-pit pumps; preliminary treatment consisting of two mechanically cleaned fine-bar screens, a four-unit vortex-cyclonic grit removal system with associated grit classifier, and an influent flow measurement via a 36-inch Parshall flume with an ultrasonic flow meter; primary treatment consisting of sedimentation in four 49,370-gallon rectangular basins and four 52,960-gallon rectangular basins; a biological treatment process consisting of a five-stage Bardenpho BNR process that includes three 250,000-gallon fermentation basins, three 333,000-gallon first anoxic reactors, 13 aeration basins or nitrification reactors (three 363,170-gallon basins, and ten 127,160-gallon basins), four 280,000-gallon second anoxic basins, and four 63,000-gallon re-aeration basins; four 100-foot-diameter secondary clarifiers; four return-activated sludge pumps; an intermediate effluent pumping station using three 60-inch-diameter Archimedes screw lifts and three centrifugal pumps; polishing filtration consisting of 12 rapid-sand, pulsed-filtration, gravity-type automatic backwash filters with a total surface area of 4,320 square feet; an effluent disinfection system using gaseous chlorination and a 315,000-gallon, dual-channel chlorine contact basin. Alum is added before the effluent reaches the polishing filters to aid in total suspended solids (TSS) removal and thereby reduce trihalomethane (THM) formation potential. Also on-site is a 5-million-gallon (MG) reclaimed water storage tank and accompanying high-service pumps.

Chlorinated effluent from the chlorine contact basin is directed to the Master Reuse System or to a 315,000-gallon dechlorination basin that uses flow-paced sulfur dioxide to eliminate the remaining chlorine residual. It then flows through a 100,000-gallon re-aeration basin and finally through a 48-inch-diameter outfall pipe that discharges to Stevenson’s Creek, 20 feet from shore.

Waste sludge from the primary clarifiers is pumped to one 930,000-gallon anaerobic digester. Waste sludge from the secondary clarifiers is pumped to two 108,000-gallon-per-day (gpd) rotary drum thickeners equipped with polymer injection, then to the anaerobic digester. The digested sludge is then directed to a 127,000-gallon sludge blend tank. The blended sludge is dewatered using two 2-meter belt filter presses.

Basis of Design and Actual Flow

Flow

The design flow for the facility is 10 MGD; the average flow for the study period was 5.48 MGD, and the maximum month flow during the study period was 6.85 MGD during September 2006.

Loadings

Plant design loadings and equipment parameters are as follows:

Average day	10 MGD
Peak day	15 MGD

Primary settling tanks: 4 each at 49,370 gallons, 4 each at 52,960 gallons
The plant operates four units regularly.

Activated-sludge

Fermentation basins: 3 each at 250,000 gal
First anoxic basin: 3 each at 333,000 gal
Aerobic basin: 3 each at 367,000 gal
Aerobic basin: 10 each at 127,000 gal
Anoxic basin: 4 each at 280,000 gal
Re-aerobic basin: 4 each at 63,000 gal

Total hydraulic retention time (HRT): 20 hours
Design mixed liquor suspended solids (MLSS): 4,000 mg/L
Return activated sludge (RAS) rate: 80–120 percent
Internal recycle rate: 400–600 percent
Food-to-microorganism (F-to-M) ratio: 0.05
Mean cells residence time (MCRT): 25–40 days

Secondary clarifier: 4 each, diameter = 100 ft at 12.5-ft depth

Surface loading rate: 318 gpd/ft^2 at average daily flow (ADF)
Detention time: 7 hours at ADF
The plant operates three units regularly.

Rapid sand, pulsed filter: 12 each, 12 ft by 30 ft, or a total of 4,320 sf

ADF capacity: 2 MGD each
Peak capacity: 28 MGD
Hydraulic loading rate: 3.8 gpm/sf at ADF
4.5 gpm/sf at peak

Sludge thickener—Carter rotary drum

Capacity: 2 each, 75 gpm
Thicken sludge: Waste-activated sludge (WAS) at 4–6 percent
Volume: 15,552 gpd

Anaerobic digester

Primary digester: Diameter = 85 ft, volume = 0.93 million gallons

Digesters—The sludge-heating system and gas-mixing system were not operational in the primary digester from October 2005 through September 2006 because the primary digester system was being rebuilt. During that period, all sludge was pumped directly to the blending tank for dewatering. The primary digester was back online in January 2007.

Dewatering—The primary sludge and WAS are blended with polymer for dewatering with an Andritz belt filter press. The cake is hauled away by truck.

Plant Parameters

Overall plant influent and effluent average results for the period October 2005 to September 2006 are shown in Table 2.

Table 2. Influent and effluent averages

Parameter (mg/L unless stated)	Average value	Maximum month	Max month vs. avg.	Maximum week	Sample method/frequency
Flow (MGD)	5.48	6.85	25%	7.62	--
Influent TP	5.0	5.53	10%	6.35	Weekly/composite
Effluent TP	0.13	0.21	62%	0.26	Weekly/composite
Influent BOD	188	234	24%	263	Daily/composite
Effluent BOD	2.3	4.1	78%	5.3	Daily/composite
Influent TSS	231	277	20%	317	Daily/composite
Effluent TSS	0.89	1.11	24%	1.6	Daily/composite
Influent NH_4-N	28.0	32	16%	34.0	Daily/composite
Effluent NH_4-N	0.036	0.045	25%	0.062	Daily/composite
Influent Total N	28.0	32	16%	34.0	Daily/composite
Effluent Total N	2.32	3.1	35%	3.75	Daily/composite

Notes:
BOD = biochemical oxygen demand
Max month vs. average = (max month – average) / average x 100
NH_4-N = ammonia measured as nitrogen
TN = total nitrogen
TP = total phosphorus
TSS = total suspended solids

Table 3 presents the plant's monthly averages for the Bardenpho process parameters.

Table 3. Monthly averages for plant process parameters

Month	MLSS (mg/L)	Sludge age (d)	HRT (hr)	Temperature (°C)
Oct 2005	3,979	51	27	29.3
Nov 2005	4,106	44	28	27.2
Dec 2005	4,181	44	30	24.6
Jan 2006	4,425	36	30	23.8
Feb 2006	4,094	27	28	23
Mar 2006	3,951	25	28	25
Apr 2006	3,857	34	27	27
May 2006	3,340	31	28	28
June 2006	3,704	41	29	30
July 2006	4,205	34	26	30
Aug 2006	3,701	37	25	31
Sep 2006	3,921	36	22	30

Notes:
HRT = hydraulic retention time
MLSS = mixed liquor suspended solids

Performance Data

Figures 3 and 4 present reliability data for TP removal. The removal is good, with the effluent TP averaging 0.13 mg/L and having a medium coefficient of variation (COV) of 40 percent. The COV is defined as the standard deviation divided by the mean, and it is a measure of the reliability of a system. The lower the COV, the less the data are spread and so the higher the reliability.

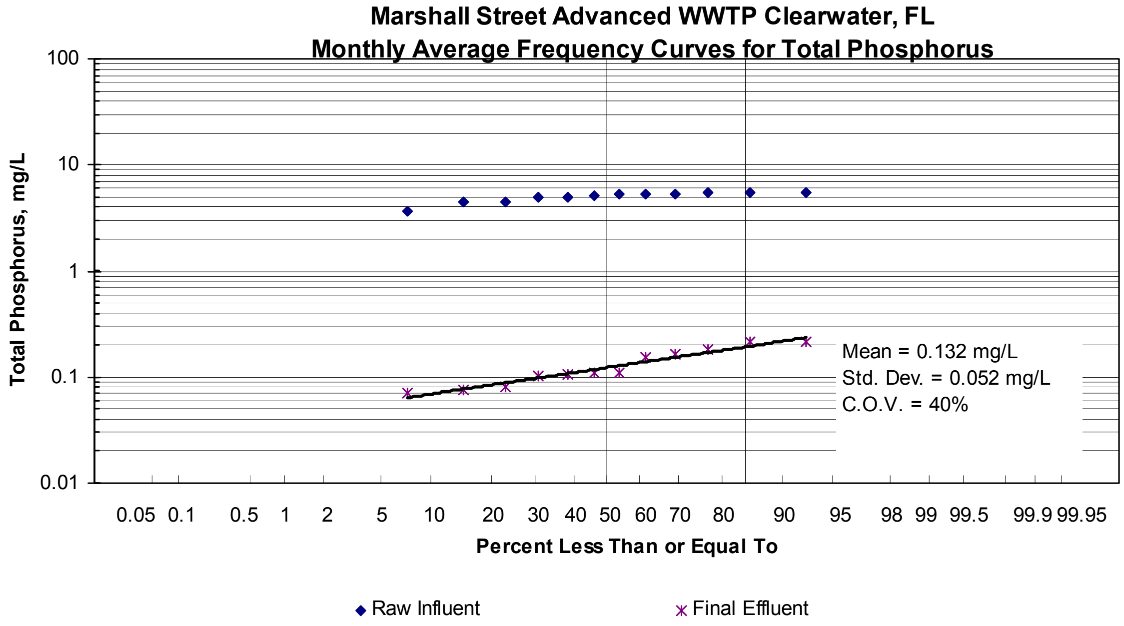

Figure 3. Monthly average frequency curves for TP.

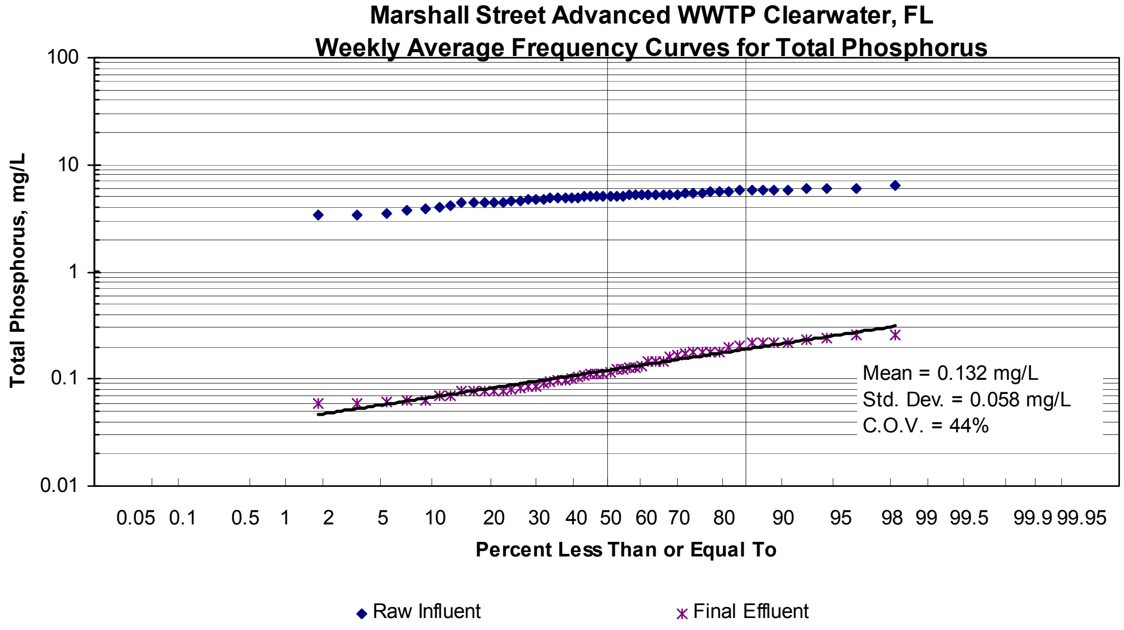

Figure 4. Weekly average frequency curves for TP.

Figures 5 and 6 present reliability data for ammonia nitrogen removal. Removal of ammonia nitrogen is very good, with a mean effluent of 0.038 mg/L and a very low COV of 18 percent.

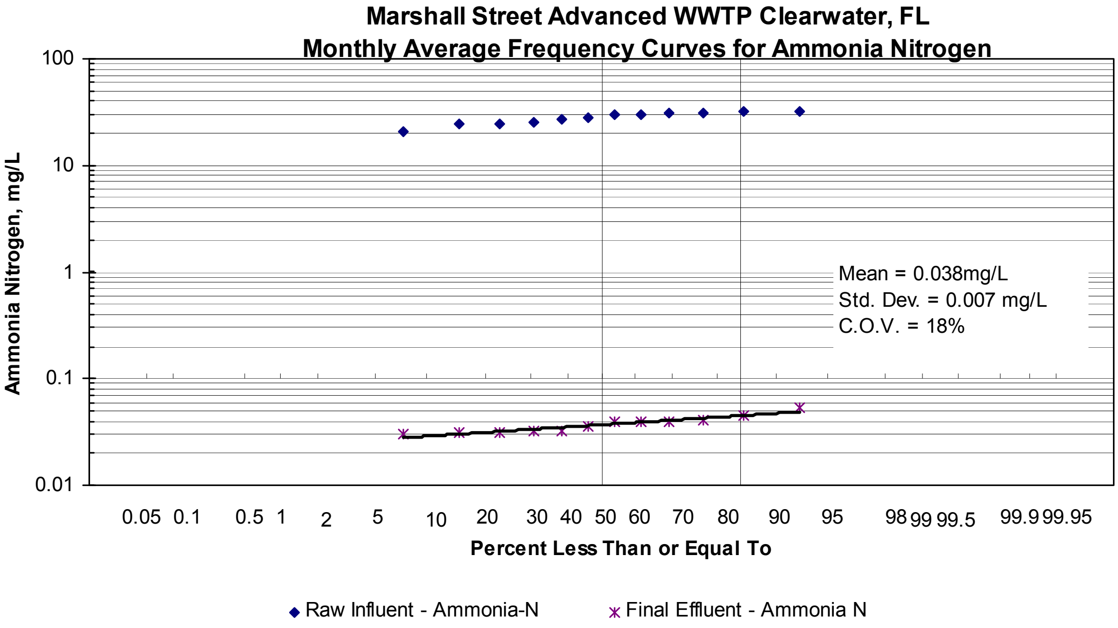

Figure 5. Monthly average frequency curves for ammonia nitrogen.

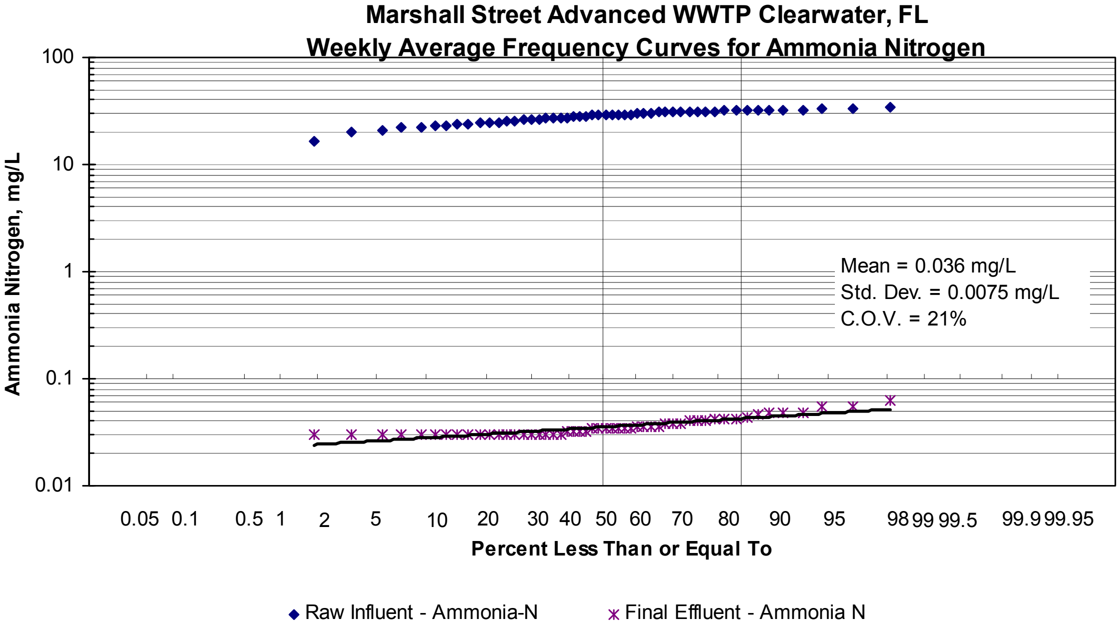

Figure 6. Weekly average frequency curves for ammonia nitrogen.

Figures 7 and 8 present reliability data for removal of TN. With the two anoxic stages, the plant gives outstanding TN removal, with effluent TN of 2.32 mg/L and a COV of 16 percent.

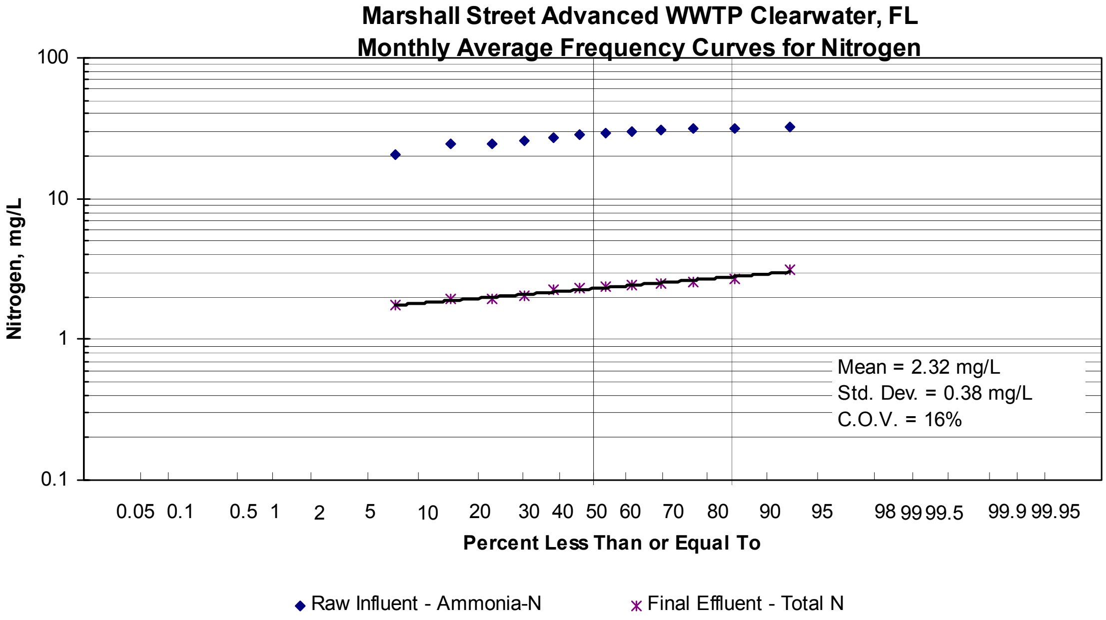

Figure 7. Monthly average frequency curves for nitrogen.

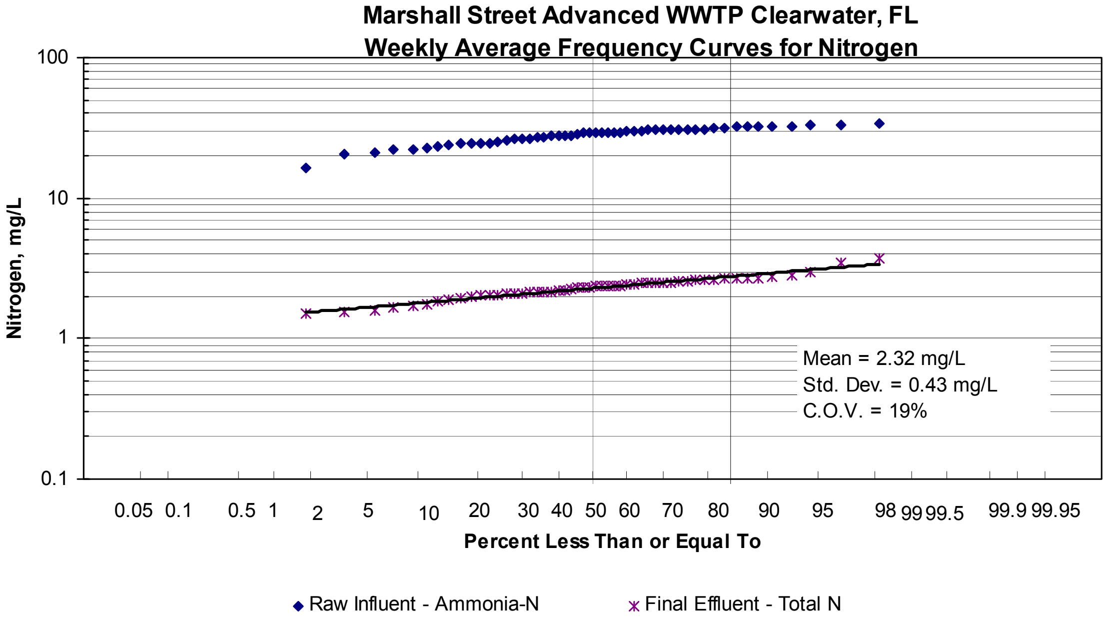

Figure 8. Weekly average frequency curves for nitrogen.

Reliability Factors

This facility's design is unique in several ways. The plant has multiple treatment processes in series to provide efficiency and reliability in meeting nitrogen and phosphorus limits. They include primary settling, a five-stage Bardenpho process for biological nitrogen and phosphorus removal, and tertiary filtration. In addition, some chemical removal of phosphorus can be obtained when alum is added before the tertiary filters for THM control. The results are excellent: the plant achieves a phosphorus mean concentration of 0.13 mg/L with a COV of 40 percent and a TN mean concentration of 2.32 mg/L with a COV of only 16 percent as a monthly average. The plant's maximum average week results were good, with the maximum average week phosphorus at 0.26 mg/L versus the weekly standard of 1.5 mg/L, the maximum average week ammonia nitrogen at 0.062 mg/L, and the maximum average week TN at 3.75 mg/L versus the weekly standard of 4.5 mg/L. These results are well within the normal range of variation from average for a wastewater treatment process, as reflected in the low to very low COVs shown in Figures 3, 5, and 7. As shown in Table 2, the fractions by which the monthly effluent maxima exceeded the corresponding annual averages (62 percent, 25 percent, and 35 percent for TP, ammonia nitrogen, and TN, respectively) were consistent with or better than the literature suggestion of 63 percent (Brandao et al. 2005). The key factors for this exceptional performance are discussed below.

Wastewater characteristics: The BOD-to-TP ratio was favorable, with an average value of 37.5, and ranged monthly between 31 and 44. A ratio of 20 is recommended in the literature (WEF and ASCE 1998). The average BOD-to-total Kjehldahl nitrogen (TKN) ratio was 6.7 and ranged monthly between 6.1 and 7.6. Both ratios are favorable for BNR. The soluble BOD-to-ammonia nitrogen ratio has been in the range of 4 to 5, less than what was originally recommended (6). It should be noted that on weekdays 160,000 gal/day of filtrate from the belt filter presses is returned to the head of the plant; this filtrate contains 51 mg/L of TP and 131 mg/L of ammonia nitrogen. These loads amount to 30 percent of the influent TP and 14 percent of influent ammonia, with the effective minimum BOD-to-TP and BOD-to-TN ratios dropping to 24 and 5.3, respectively. The soluble BOD-to-ammonia nitrogen ratio similarly drops to 3.7. Despite these recycle stream loads and the low BOD-to-ammonia nitrogen ratio, no adverse effect was reported under the operating parameters developed at this facility.

Primary settling tanks: The plant regularly operates four tanks out of the eight available, and the efficiencies in removal are typical—30 percent in BOD and 50 percent in TSS.

Activated sludge: The five-stage Bardenpho process at the facility is a typical design. It includes a fermentation zone, followed by the first anoxic and aerobic zones in series, a second anoxic zone, and the re-aeration zone. The typical internal recirculation of MLSS to the first anoxic zone from the second aerobic zone is five times the influent flow rate. Some unique features of this process are two separate anoxic zones, each with long detention times of approximately 1.5 hours, long sludge age ranging between 30 and 50 days, and high water temperature.

Phosphorus removal far exceeds the permit requirement with good reliability and is achieved by two processes: first by the Bardenpho process as the primary process, then later as a side benefit to alum addition, which is done primarily to reduce TSS and so reduce potential THM formation. The Marshall Street WRF has to meet a limit on dichlorobromomethane of 22 micrograms per liter (µg/L) and a dibromochloromethane limit of 34 µg/L to meet Florida's state requirements for water reuse. The typical dosage of alum is 27 mg/L, or 2.4 mg/L as aluminum (Al). This dosage is equivalent to an Al-to-TP ratio of 1.6 on a molar basis in the plant influent. For the effluent concentration the plant produces, this ratio is considered low for a strictly chemical removal process.

Nitrogen removal has been excellent with good reliability. No external carbon source is used. The use of two anoxic zones with an internal recirculation flow rate of five times the influent flow rate has been found to be sufficient to produce low nitrogen concentrations (WEF and ASCE 1998). It is also noteworthy that the plant maintains a sludge blanket in the secondary clarifiers. The depth ranges between 2 and 3 feet and is a part of the TN removal strategy and the biological phosphorus removal strategy.

Another key operational factor is the automated process control system, which uses Chemscan and supervisory control and data acquisition (SCADA). These programs monitor online at the second anoxic zone nitrate-nitrogen, dissolved oxygen (DO), oxidation-reduction potential (ORP), and ortho-phosphorus to optimize nitrogen removal. Table 4 lists the sensors used at the Marshall Street facility. The minimum ORP is set at –60 millivolts (mV), and the nitrate-nitrogen is set at a minimum of 0.5 mg/L. The DO is adjusted on the basis of these two parameters. In addition, the system monitors MLSS and the sludge blanket in the secondary clarifiers. The plant also has monitors for turbidity, in accordance with the permit, and conductivity, to monitor for salts that could intrude by means of seawater and adversely affect irrigation reuse. All the automation and controls have contributed to an efficient phosphorus removal and full denitrification with good reliability.

Table 4. Probe and sensor suppliers

Parameter	Supplier(s)
Dissolved oxygen	Hach, Royce
MLSS	Hach
Nitrate-nitrogen	Chemscan
Ammonia nitrogen	Chemscan
Clarifier sludge blanket depth	Hach, Royce
pH	Hach
Oxidation-reduction potential (ORP)	Hach
Ortho-phosphorus	Chemscan
Turbidity	Hach

In addition, this plant has the flexibility of operating as a four-stage Bardenpho process, thereby providing additional tank volume dedicated to nitrogen removal. Under this mode of operation, the phosphorus removal is achieved primarily by alum addition.

Secondary clarifiers: The plant regularly operates three out of four units at the current flow. One practice to note is the maintenance of the sludge blanket at between 2 and 3 feet, which is monitored with the new blanket monitors installed in 2002.

Tertiary filter: The tertiary filter is an original Zimpro filter with air and water backwash provisions. The system is effective in suspended solids removal: the effluent TSS averages 1 mg/L or less. This, in turn, is a key to achieving the low phosphorus concentration in the final effluent.

Recycle flows from dewatering and thickening go back to the primary clarifier influent. The returns are controlled to flow uniformly around the clock and avoid a shock loading to the treatment processes. No adverse impact has been observed under this practice at this facility.

Another key parameter to note is the long sludge age maintained at this plant. Because of this long sludge age at warm temperature ranges, a sludge yield of around 0.25–0.4 lb volatile suspended solids (VSS) per lb of BOD removed has been reported. This is consistent with Manual of Practice No. 8 (WEF and ASCE 1998). This low yield naturally contributes to a low cost in sludge handling.

Costs

Capital Costs

The main upgrade of the plant for BNR occurred in 1988 when the basins were reconfigured for the five-stage Bardenpho process. The upgrade then cost $16.8 million, which was updated to $29.5 million in 2007 dollars using the *Engineering News-Record* (USDA 2007). The upgrade included additional tanks or dividing walls, mixers, pumps, blowers/aerators and tertiary filtration.

It was assumed that 17 percent of the upgrade was attributed to phosphorus removal, while 63 percent of the upgrade was for nitrogen removal. This allocation was done in consultation with plant personnel and was based on the fraction of the secondary system volume that could be attributed to phosphorus or nitrogen removal. Specifically, all anaerobic tank volume plus 10 percent of the volume of the aerobic tanks (based on oxygen usage) was attributed to phosphorus removal, while all anoxic tank volume plus 50 percent of the aerobic tanks (based on oxygen usage) was attributed to nitrogen removal. The balance of the upgrade was attributed to BOD removal or other activities required by permit. The tertiary filters were installed to meet the requirements for surface water discharge under reuse rule

62-610 in Florida. This meant that the capital expenditure in 2007 dollars that was attributed to phosphorus removal was $5.02 million. The annualized capital charge (20 years at 6 percent) was $438,000 for phosphorus removal.

The capital expenditure attributed to nitrogen removal was $10.6 million in 2007 dollars. The annualized capital charge (20 years at 6 percent) was $1.6 million for nitrogen removal.

The total capital expenditure attributed to BNR was $29.5 million in 2007 dollars. For the 10-MGD facility, the capital expenditure per gallon of BNR treatment capacity was $2.95.

Operation and Maintenance Costs

In all case studies prepared for this document, the O&M costs considered were for electricity, chemicals, and sludge disposal. Labor costs for O&M were specifically excluded for three reasons:

1. Labor costs are highly sensitive to local conditions, such as the prevailing wage rate, the relative strength of the local economy, the presence of unions, and other factors; thus, they would only confound comparison of the inherent cost of various technologies.

2. For most processes, the incremental extra labor involved in carrying out nutrient removal is recognized but not significant in view of the automatic controls and SCADA system that accompany most upgrades.

3. Most facilities were unable to break down which extra personnel were employed because of nutrient removal and related overtime costs, making labor cost development difficult.

CAPDETWorks was used to provide a relative comparison of labor costs compared to power costs. CAPDETWorks is a software package developed by Hydromantis Corporation (Hamilton, Ontario, Canada). It is used to estimate conceptual capital and operating cost estimates for wastewater treatment facilities. It is based on work originally done by EPA and the U.S. Army Corps of Engineers. Two flow scenarios were run for a model consisting of a five-stage Bardenpho reactor, a secondary clarifier, a tertiary filter, and an anaerobic digester: (1) 5.5 MGD to mimic the current flow at the plant and (2) 10 MGD to match the design flow. For 5.5 MGD, the CAPDET electrical cost estimate using the plant's overall average rate of $0.11 per kilowatt-hour (kWh) was $960,000, while the O&M labor cost at an average regional rate of $35/hour was $540,000. For a 10-MGD facility, the CAPDET electrical cost estimate was $1.7 million, while the labor cost was $680,000. By comparison, as shown below, the Marshall Street facility's electrical cost for similar equipment at an average flow of 5.5 MGD was $840,000, including electrical costs for BOD removal.

The plant uses both biological phosphorus and nitrogen removal, with minimal use of alum and no use of supplemental carbon sources. The plant could use minimal chemicals because the ratios of influent BOD to TP and influent BOD to TN were both very high (37.6 and 6.7, respectively). This means that costs for nutrient removal are essentially all electrical. A summary of the electrical use calculations is provided in the Attachment. The specific electrical usage for phosphorus removal was 931,000 kWh per year (kWh/yr). The average electrical rate for the plant was $0.11/kWh, and it was based on the cost per kWh plus a demand charge plus a Florida-required fuel surcharge. When that rate was applied, the cost for phosphorus removal was $102,400 for the year. The total electrical usage for nitrogen removal was 4,620,000 kWh/yr, or $509,000. The electrical usage for BOD removal in the system was 2,091,000 kWh/yr, or $230,000.

Alum is applied as an effluent-polishing step primarily for reducing THM formation potential; however, some phosphorus removal does occur with alum addition. The total cost of alum used over the evaluation period was $74,000. On the basis of the dosage of alum and the possible removal that could occur, it was assumed that 10 percent of the alum could be attributed to phosphorus removal; the chemical cost for phosphorus removal was therefore $7,400. All the alum added (2.4 mg/L as Al) was assumed to convert to aluminum hydroxide sludge; at the average flow of 5.48 MGD, this was 317 lb of aluminum sludge per day, or 58 dry tons/yr. Assuming that phosphorus removal accounted for 10 percent of the sludge and using the plant's cost of sludge disposal of $253/dry ton, the chemical sludge cost for phosphorus removal was $1,463.

Unit Costs for Nitrogen and Phosphorus Removal

During the evaluation period, the plant removed 81,200 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal was $1.37 per pound, while the annualized unit capital cost was $5.39/lb of phosphorus removed. At design flow, the annualized capital would drop to $2.95/lb of phosphorus removed.

During the evaluation period, the plant removed 428,000 lb of TN. With the results above, the unit O&M cost for TN removal was $1.18/lb, while the annualized unit capital cost is $3.79/lb of nitrogen removed. At design flow, the annualized capital cost would drop to $2.07/lb of TN removed.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the annualized unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal was $6.76/lb phosphorus removed, while the life-cycle cost for TN removal was $4.97/lb nitrogen removed, all at current flows. At design flows, assuming the O&M costs increase proportionally to flow and loadings, the life-cycle costs would be $4.32/lb of phosphorus removed and $3.25/lb of TN removed.

Assessment of magnitude of costs: The capital cost of $2.95/gpd capacity is relatively high, but the O&M costs remain low. One of the key factors is that no methanol is purchased because of the use of the incoming carbon source for both nitrogen and phosphorus removal with the five-stage Bardenpho process.

Discussion

Reliability factors: The treatment processes at the Marshall Street plant represent a traditional layout for the original five-stage Bardenpho process for both biological nitrogen and phosphorus removal—one anaerobic zone, two anoxic zones with a high rate of internal recirculation, an aeration zone in between, and the final re-aeration zone. This is accomplished with a conservative design basis—a long HRT, a long sludge age, and a low clarifier loading rate in a warm-temperature region. Another key is the automated controls the plant personnel use, which are based on online monitoring with multiple sensors and process control parameters for the Bardenpho process. In addition, good primary settling tanks and efficient tertiary filters added reliability along with alum addition for effluent THM reduction. This process as operated by the plant personnel has proven to be efficient and reliable in meeting the permit limits of 3 mg/L for nitrogen and performing significantly better than the limit of 0.2 mg/L in phosphorus.

Cost factors: The costs are relatively high for capital but low for O&M. This plant was designed with conservative design parameters, at $2.95/gpd capacity. The O&M costs are low at $1.37/lb of phosphorus removed and $1.18/lb of TN removed. The main reasons for these low costs are efficient operation of the biological processes and no need for an external carbon source (e.g., methanol). Even though the power cost in Florida, compared to that of other states, is high at $0.11/ kWh, the overall O&M cost is relatively low. In addition, the alum addition is at a reduced dosage and thus the cost impact is low because the Bardenpho process removes a significant amount of phosphorus biologically. All these costs are based on the plant's current flow. As the plant flow increases to the full design loadings, these unit costs would be expected to decrease.

Summary

The Marshall Street WRF is an advanced wastewater treatment plant with a five-stage Bardenpho process that meets the effluent discharge limit for nitrogen and exceeds that for phosphorus. The reliability has been excellent in achieving low concentrations—0.13 mg/L in phosphorus with a COV of 40 percent and 2.32 mg/L in nitrogen with a COV of 16 percent monthly average. The cost for this facility is considered high with a capital cost at $2.95/gpd capacity, but the O&M costs are low. The unit costs are low at $6.76/lb of phosphorus removed and $4.97/lb of TN removed.

Key contributing factors for reliability include favorable wastewater characteristics, conservative design with multiple processes in series, good operating procedures for the Bardenpho process developed by the plant personnel, and automation with online sensors and control devices.

Key contributing factors to facility costs include a conservative design originally, an efficient operation without an external carbon source, and optimization of energy and chemical usage, while minimizing sludge production from the biological process.

Acknowledgments

The authors are grateful for the significant assistance and guidance provided by John Milligan, superintendent of the Wastewater Environmental Technologies Division, Clearwater; Tom Nietzel, coordinator of the Wastewater Environmental Technologies Division, Clearwater; and Jeff Borden, chief operator of the Marshall Street WRF. This case study would not have been possible without their prompt response with well-deserved pride in the facility and its operation. EPA thanks Clearwater, Florida, for participating in this case study.

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Attachment: Electrical Use and Chemical Costs

	Horse-power	#	kW power draw	hours/ day	kWh draw/day	kWh draw/year	% BOD	%P	%N	Usage for BOD	Usage for P	Usage for N
Ferment basin mixers	10	6	44.76	24	1,074.24	392,097.6	0%	100%	0%	0	392,097.6	0
1st anoxic mixers	7.5	9	50.355	24	1,208.52	441,109.8	0%	0%	100%	0	0	441,109.8
Aerator	400	2	596.8	24	14,323.2	5,227,968	40%	10%	50%	2,091,000	522,796.8	2,613,984
Pumps—internal recycle	50	3	111.9	24	2,685.6	980,244	0%	0%	100%	0	0	980,244
2nd anoxic mixers	7.5	12	67.14	24	1,611.36	588,146.4	0%	0%	100%	0	0	588,146.4
Filter lift pumps	50	1	37.3	24	895.2	326,748	0%	5%	0%	0	16,337.4	0
Total draw kWh/yr						7,629,566				2,091,187	931,231.8	4,623,484

Alum use	$74,000
% for P	10
Alum cost for P	$7,400

Noman M. Cole, Jr., Pollution Control Plant
Fairfax County, Virginia

Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

This facility was selected for as case study because it employs a step-feed activated-sludge strategy with tertiary filters and a ferric chloride feed.

The Noman M. Cole, Jr., Pollution Control Plant serves the area of Fairfax County, Virginia, in the Washington, D.C., metropolitan area. The plant was originally placed in operation in 1970. The average wastewater treatment capacity of the plant was 18 million gallons per day (MGD) when commissioned; this has risen to 67 MGD after a series of successful expansions. Biological nutrient removal (BNR) was added in 2002 as part of a 13-MGD expansion.

The Virginia Pollutant Discharge Elimination System (VPDES) permit limits for the Noman M. Cole, Jr., Pollution Control Plant are shown in Table 1.

Table 1. VPDES permit limits

Parameter	Monthly average (mg/L)	Monthly average (lb/day)	Weekly average (mg/L)	Weekly average (lb/day)
CBOD	5	2,790	8	4,464
TSS	6	3,348	9	5,020
Ammonia-N (April–Oct)	1.0	559	1.5	836
Ammonia-N (Nov–Mar)	2.2	--	2.7	--
Total N	Report	--	Report	--
Total P	0.18	101	0.27	150

Notes:
CBOD = carbonaceous biochemical oxygen demand
N = nitrogen
P = phosphorus
TSS = total suspended solids

Treatment Processes

The facility uses a step-feed strategy to distribute organic matter throughout the biological treatment basins. Following primary settling, the flow goes to a set of nine aeration basins that are operated in anaerobic, aerobic, or anoxic modes. The activated-sludge process was designed for a normal detention time of 8.9 hours with up to five feed points into the basin. Feed is typically distributed to three anaerobic or anoxic points in the system. Polymer can be added to aid secondary clarification. The facility uses ferric chloride and polymer to polish the secondary effluent, primary-to-tertiary clarification, and filtration. The final effluent is chlorinated/dechlorinated before discharge to Pohick Creek, a tributary to the Potomac River.

The primary sludge is fermented in the gravity thickeners at a sludge residence time (SRT) of 3 days and a hydraulic retention time (HRT) of less than 24 hours. The secondary sludge is thickened at the dissolved air flotation (DAF) units. The fermented primary sludge and thickened secondary sludge are mixed together for dewatering by centrifuge, followed by incineration. Lime can be added to the dewatering process to minimize recycle loads of nutrients.

Figure 1 shows the plant’s flow schematic. The secondary system consists of nine parallel aeration basins—six small (1.67 million gallon [MG] total volume each) and three large (4.89 MG total volume each). Figure 2 shows how the step-feed works in the larger basins. The feed can be provided at five anoxic zones through each basin, although in practice only four (A, C, D, and E) receive feed. The smaller basins have three points for step-feeding primary effluent. Under normal circumstances, the flow split between zones A, B, and C in the smaller basins is 40 percent, 40 percent, and 20 percent, respectively, while the larger basins, zones A, C, D, and E, each get 25 percent of the flow. Other design information on the facility is provided in Table 2 and the attachment.

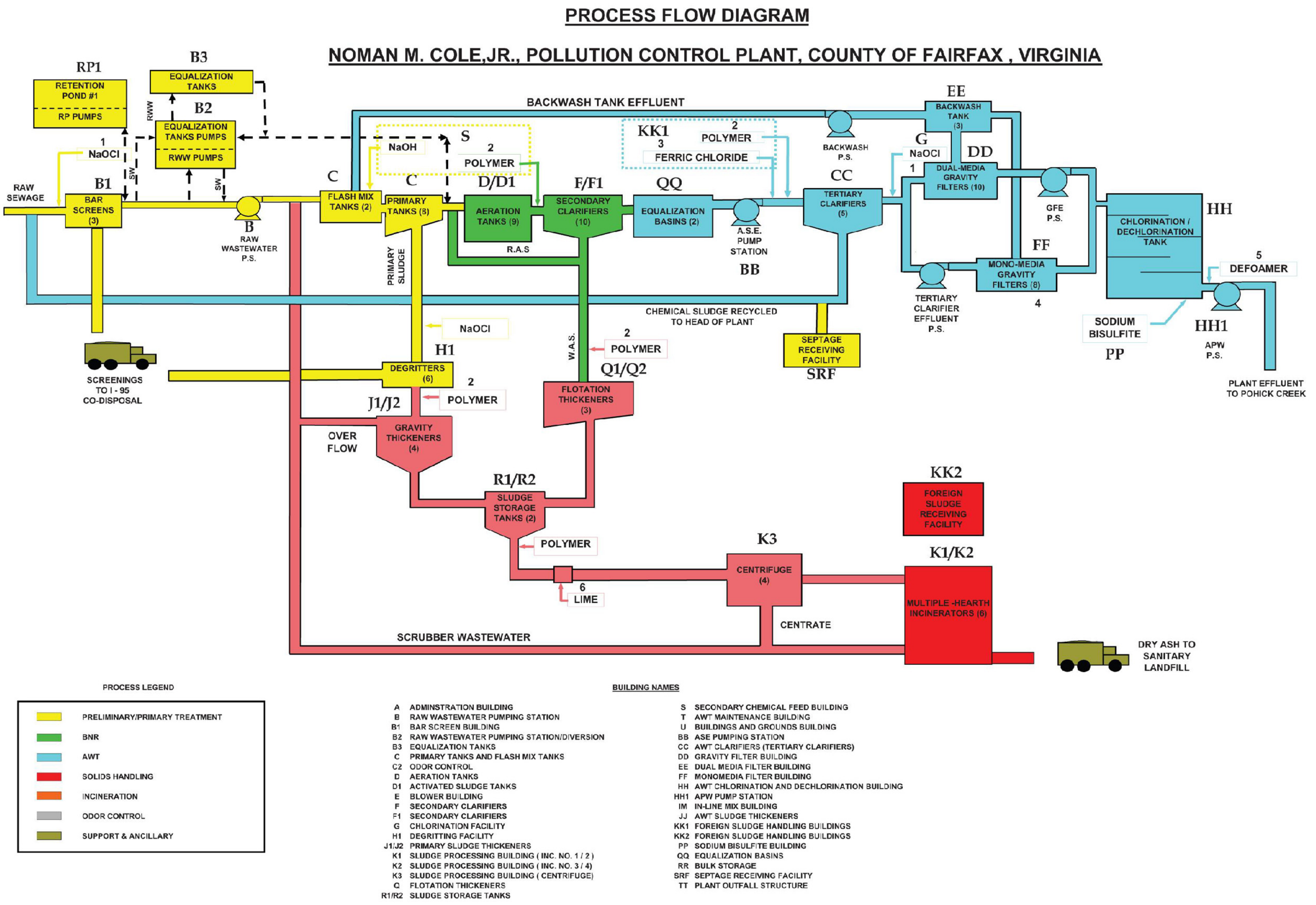

Figure 1. Noman M. Cole, Jr., Pollution Control Plant process flow.

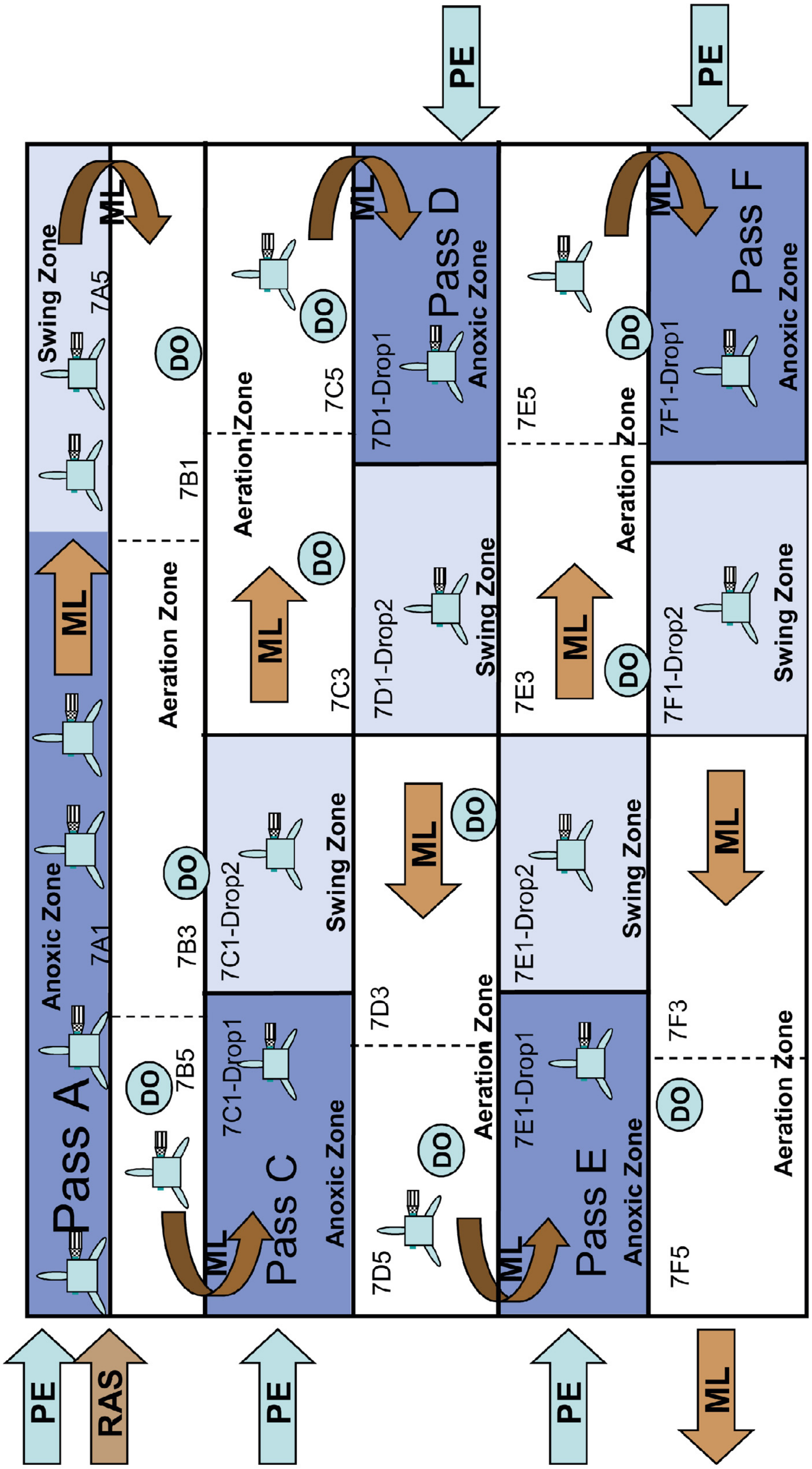

Figure 2. Noman M. Cole, Jr., Pollution Control Plant step-feed detail.

Table 2. Facility design data

Secondary tanks	**Tanks 1–6**	**Tanks 7–9**
Volume	1.67 MG	4.89 MG
Anoxic volume	3.5–5 MG	5.1–7.3 MG
HRT (average)	8.9 hr	8.9 hrs
SRT at maximum month loading (last MLSS 4,400 mg/L)	18 days	18 days
Gravity thickeners (2)		
Volume	0.146 MG each	
SRT	3 days	
HRT	> 12 hours	
Tertiary clarifier		
Diameter	152 ft	
Hydraulic loading rate	735 gpd/sf (average flow)	
Tertiary filters	Monomedia	Gravity filters
Number	8	10
Media type	Anthracite	Garnet/sand/anthracite
Depth	5 ft	2.25 ft
Design loading rate, gpm/sf	2.9	2.6
Dimensions	30 ft x 17 ft x 2 cells	30 ft x 30 ft x 2 cells

Notes:
gpd/fs = gallons per day per square foot
HRT = hydraulic retention time
MG = million gallons
MLSS = mixed liquor suspended solids
SRT = solids retention time

Plant Parameters

Overall plant influent and effluent average results for the 2006 calendar year are shown in Table 3.

Table 3. Influent and effluent averages

Parameter (mg/L unless stated)	Average value	Maximum month	Max month vs. Avg.	Maximum week	Sample method/frequency
Flow (MGD)	47.4	51.4	8%	54.4	Daily
Influent TP (mg/L)	6.39	7.06	10%	8.16	Composite/daily
Effluent TP (mg/L)	0.09	0.12	33%	0.16	Composite/daily
Influent BOD (mg/L)	189	205	8%	305	Composite/daily
Effluent BOD (mg/L)	2.0	2.0	0%	2.0	Composite/daily
Influent TSS (mg/L)	225	253	12%	353	Composite/daily
Effluent TSS (mg/L)	1.0	2.2	120%	3.06	Composite/daily
Influent NH_4-N (mg/L)	18.9	22.5	19%	24.8	Composite/weekly
Effluent NH_4-N (mg/L)	0.12	0.15	25%	0.29	Composite/weekly
Influent TKN (mg/L)	34.6	40.4	17%	48.1	Composite/weekly
Effluent TKN (mg/L)	0.9	1.12	26%	1.6	Composite/weekly
Effluent NO_3/NO_2 (mg/L)	4.35	5.03	16%	6.41	Composite/weekly

Notes:
TP = total phosphorus
BOD = biochemical oxygen demand
TSS = total suspended solids
TKN = total Kjeldahl nitrogen
NH_4-N = ammonia measured as nitrogen
NO_3 = nitrate
NO_2 = nitrite
NH_4-N = ammonia measured as nitrogen
NO_3 = nitrate
NO_2 = nitrite

Table 4 presents plant monthly average plant process parameters.

Table 4. Monthly averages for plant process parameters

Month	MLSS[a] (mg/L)	Sludge age/mean cell residence time (d)	HRT (hr)	Temperature (°C)
Jan 2006	3,626	18	9.2	17.9
Feb 2006	3,267	19	9	15.3
Mar 2006	3,390	19	9.2	17.4
Apr 2006	2,851	19	10	19.5
May 2006	3,142	18	9.8	20.6
June 2006	2,784	18	8.8	22.5
July 2006	2,383	17	8.3	23.8
Aug 2006	3,139	17	8	25.7
Sept 2006	3,192	17	7.8	25.4
Oct 2006	2,922	16	8.2	23.5
Nov 2006	2,403	16	9.4	21.2
Dec 2006	2,852	18	10	19.4

[a] MLSS is the combined average of last pass (C-PASS for AST 1-6, F-PASS for AST 7-9).

Table 5. Monthly average BOD/TP and BOD/TKN ratios

Month	Influent BOD/TP	Primary effluent BOD/TP	Influent BOD/TKN
Jan 2006	33.1	29.8	6.1
Feb 2006	33.7	29.5	5.4
Mar 2006	28.3	27.4	5.3
Apr 2006	28.2	27.1	5.5
May 2006	27.2	26.8	4.7
June 2006	28.9	24.4	5.5
July 2006	28.8	26.1	5.9
Aug 2006	29.4	28.1	4.6
Sept 2006	31.5	32.4	5.0
Oct 2006	33.5	33.8	4.9
Nov 2006	32.2	32.0	5.4
Dec 2006	28.2	26.3	5.4

Performance Data

This section provides information about the operational performance of nutrient removal at the plant. Figures 3 and 4 present the facility's 2006 monthly and weekly reliability data for phosphorus removal. The average phosphorus effluent concentration was 0.09 mg/L with a coefficient of variation (COV) of 21 percent on a monthly average basis. The COV is defined as the standard deviation divided by the mean and is a measure of the reliability of a system. The lower the COV, the less the data are spread and the higher the reliability. The phosphorus concentration exhibited a low COV of 28 percent for the weekly averages. The plant's performance in 2006 was excellent: the weekly average never exceeded even the monthly limit. The secondary effluent exhibited an average of 0.7 mg/L for the year. These figures demonstrate that both the tertiary clarifier with chemical addition and tertiary filters are key factors in meeting the permit limit at all times. Note also that the primary influent contains higher total phosphorus (TP) than the raw influent because of internal recirculation flows at the facility.

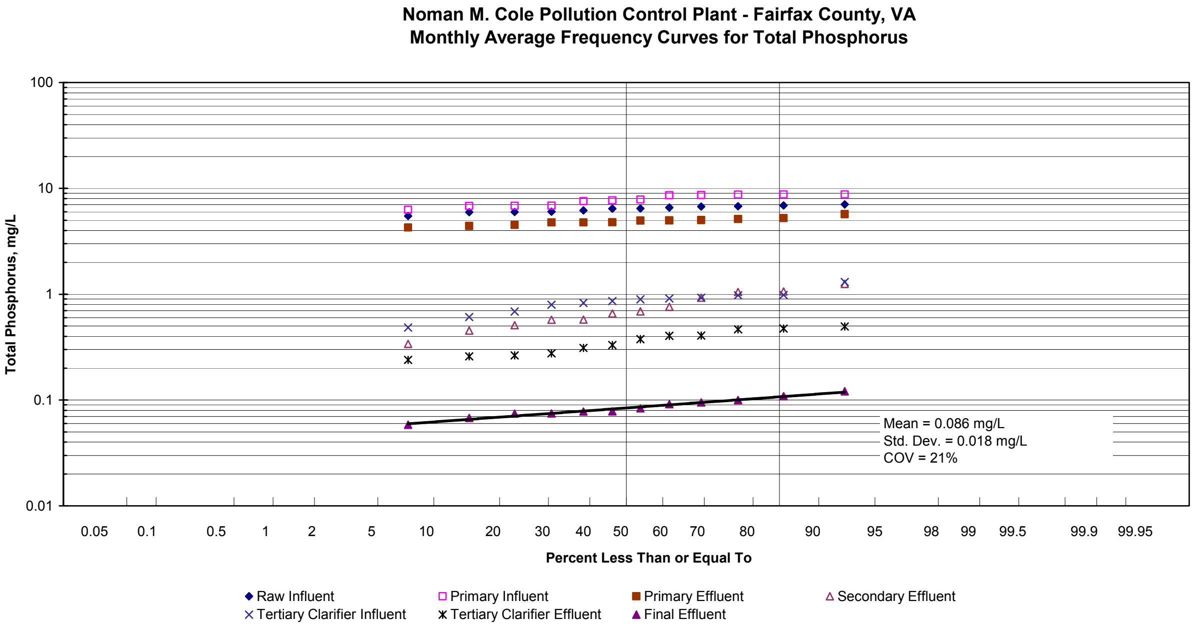

Figure 3. Monthly average frequency curves for TP.

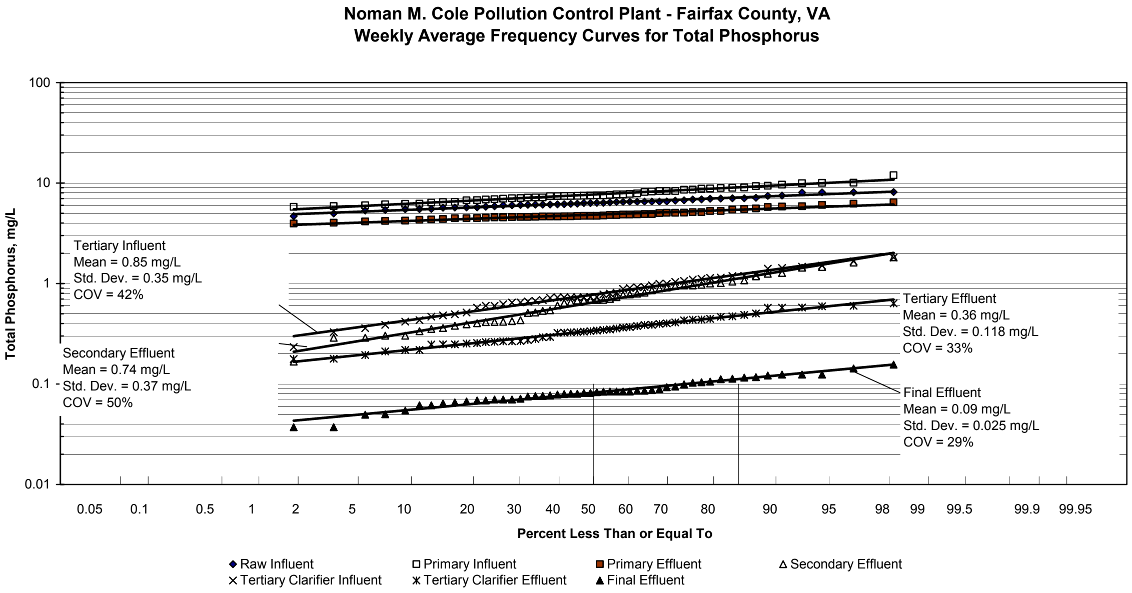

Figure 4. Weekly average frequency curves for TP.

Figures 5 and 6 present the 2006 monthly and weekly reliability data for ammonia nitrogen removal. The weekly effluent ammonia concentration averaged 0.12 mg/L, with a standard deviation of 0.035, giving a COV of 29 percent. The plant’s performance in 2006 was excellent: the weekly average never exceeded 0.3 mg/L, compared to the monthly standard of 1 mg/L during the summer months.

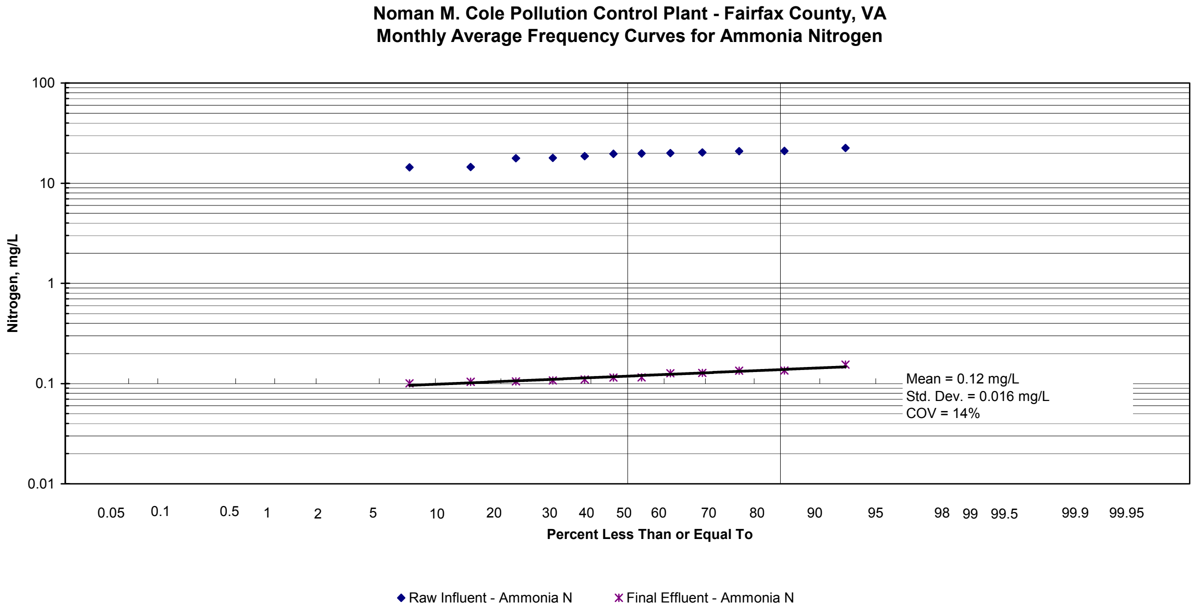

Figure 5. Monthly average frequency curves for ammonia nitrogen.

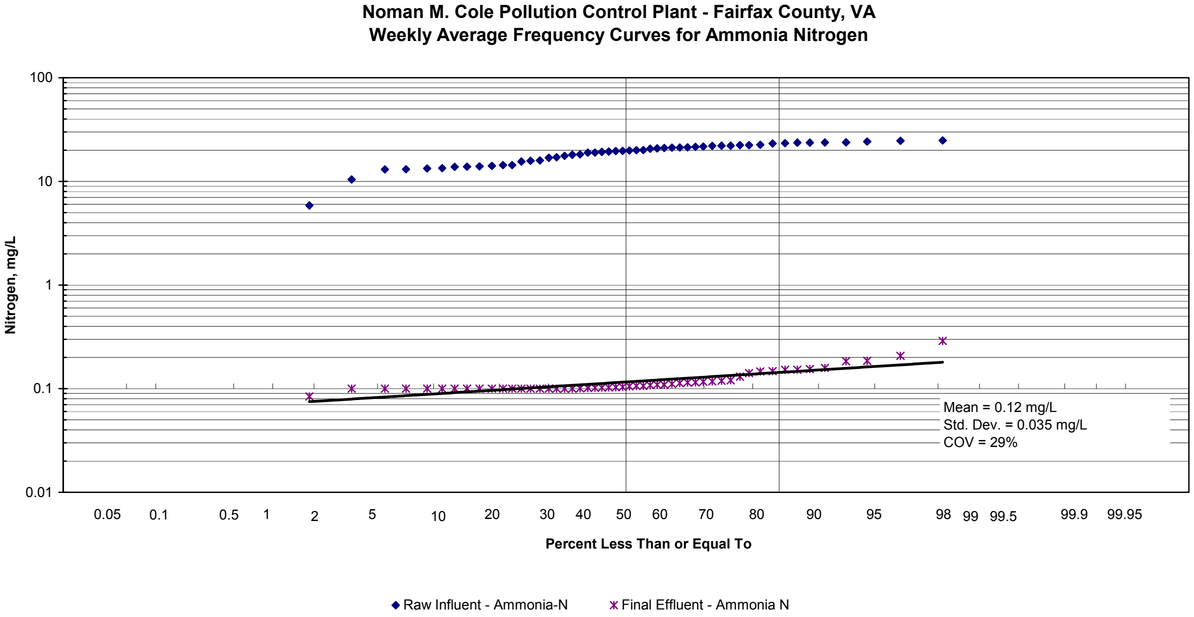

Figure 6. Weekly average frequency curves for ammonia nitrogen.

Figures 7 and 8 present the 2006 monthly and weekly reliability data for total nitrogen (TN) removal. The weekly effluent TN averaged 5.12 mg/L, with a standard deviation of 1.02 mg/L, giving a COV of 20 percent.

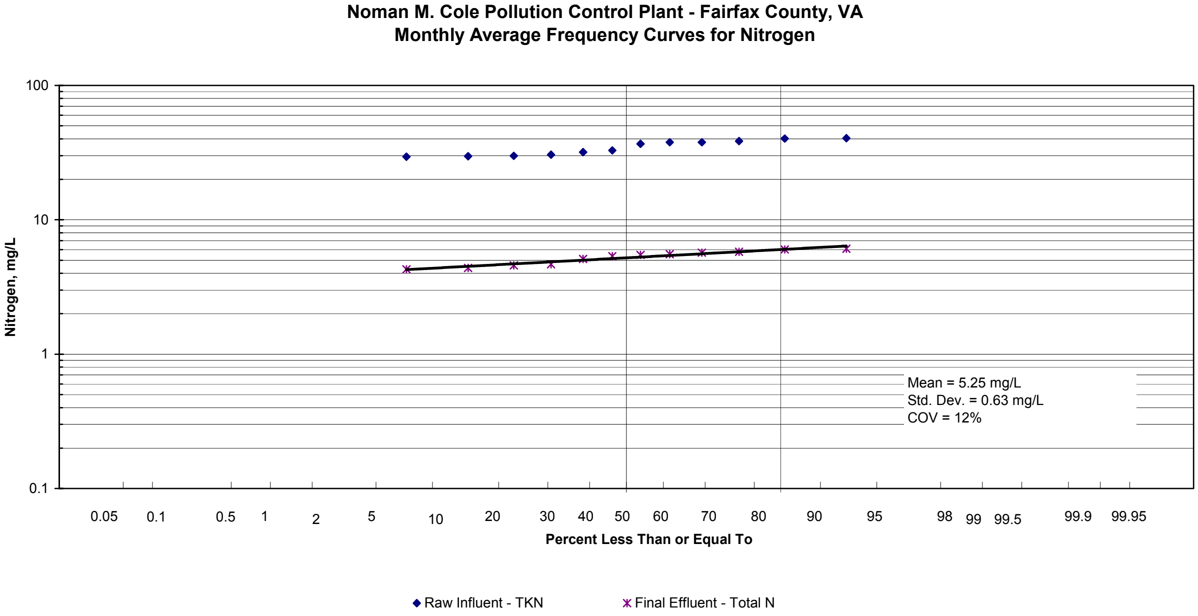

Figure 7. Monthly average frequency curves for nitrogen.

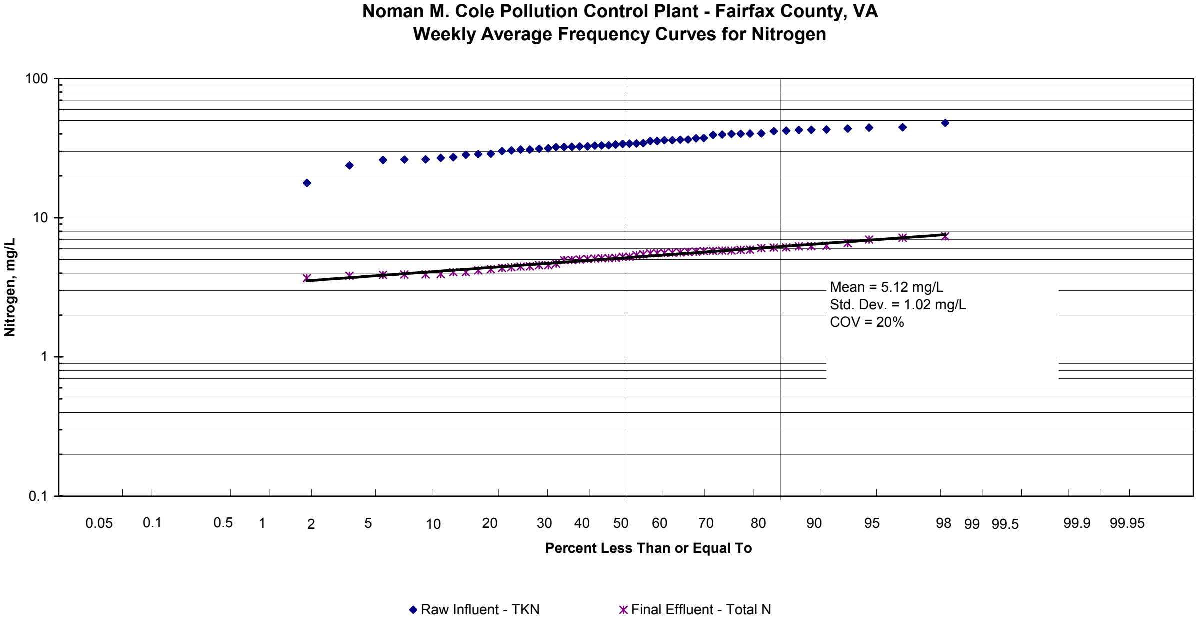

Figure 8. Weekly average frequency curves for nitrogen.

Reliability Factors

The plant has a permit limit for phosphorus of 0.18 mg/L as a year-round monthly average and monthly average ammonia nitrogen limits of 1.0 mg/L for the summer months and 2.2 mg/L for the winter months. The plant personnel have a policy of operating the plant such that these limits are seldom even approached much less exceeded. The overall reliability was good, with COVs of 21 percent for TP at the mean concentration of 0.09 mg/L, 14 percent for ammonia nitrogen at the mean concentration of 0.12 mg/L, and 12 percent for TN at the mean concentration of 5.25 mg/L for the monthly average.

A key factor in the high reliability of this step-feed plant is the care that operating staff take to ensure that any process problems do not become uncontrollable. Attention to operating details and taking appropriate and timely actions in response to plant performance data go a long way toward attaining good plant performance. It has been found that encouraging operating staff to use field test kits (e.g., Hach kits) to determine nitrogen and phosphorus concentrations provides a number of benefits, including allowing staff to take immediate action to fine-tune chemical addition and any adjustments to the biological system rather than waiting for laboratory results. It also results in a sense of ownership of the test data because they did the tests themselves. The plant has an operator for the secondary system on duty 24 hours a day, 7 days a week, and there is daily interaction between operators and engineers to review the process. A BioWin model is also used to run scenarios.

Phosphorus removal is achieved in three steps—biological removal in activated sludge, chemical removal in a tertiary clarifier, and then tertiary filters. McGrath et al. (2005) reported that biological phosphorus removal occurs when low nitrates cause the first unaerated zone to become anaerobic. Thus, the amount of nitrate returns through return activated sludge could directly affect biological removal. When nitrate levels go above 6 mg/L in the secondary effluent, biological phosphorus removal is greatly reduced. This is why the main removal mechanism for phosphorus is chemical addition followed by tertiary clarification and filtration. This sequence of operations ensures sufficient phosphorus removal, especially with chemical addition under close control by plant operators. Under current operating conditions, the operators treat any removal of phosphorus in the biological system as a bonus.

Primary sludge was fermented in gravity thickeners with an SRT of 3 days and an HRT of less than 24 hours. The volatile fatty acids (VFA) production was equivalent to 10 mg/L in chemical oxygen demand (COD) in the primary effluent, and the VFAs consisted of 33 percent acetic acid, 49 percent propionic acid, and 18 percent others (McGrath et al. 2004).

The secondary sludge was thickened at the DAF unit, thereby preventing release of phosphorus and ammonia.

Using step-feed is the primary means by which nitrogen removal through multiple anoxic zones is achieved. In the smaller biological reactors, the flow is split among three passes on a 40 percent, 40 percent, and 20 percent basis, with each pass having an anoxic zone and an oxic zone. Thus, the flow entering the first pass goes through three sets of anoxic/oxic zones, while the flow from the second pass goes through two sets of zones. In the larger basins, feed is sent to four points on the basis of 25 percent each. The system offers reliable operation because it allows using the carbon in the wastewater for denitrification rather than having to add a supplemental carbon source like methanol. Avoiding the need for supplemental carbon ensures a more economical operation because there is no need for additional feed pumps, storage tanks, and distribution and control equipment or additional sludge handling.

Recycle loads went to the primary influent, and they averaged 10 percent in biochemical oxygen demand (BOD), 19 percent in total suspended total suspended solids (TSS) and 23 percent in TP. All processes were sized to treat these recycle flows, including lime addition to the dewatering to minimize recycle loads.

The wet-weather operation included four distinct steps—retention basin (5.7 MG) first, then equalization at the headworks (4 MG), step-feed activated sludge, and finally equalization of secondary effluent (13.2 MG). The step-feed makes the process more stable than that at other plants. The holding capacity at the headworks area was equivalent to 15 percent of the design flow rate, a significant factor for good reliability.

Finally, the reliability of the plant is enhanced by a well-designed and maintained control and monitoring system, supplemented by field testing. The dissolved oxygen probes are frequently calibrated and maintained, and the plant's supervisory control and data acquisition (SCADA) system is well designed. An instrument technician is available on-site and ensures proper maintenance at this facility.

Costs

Capital Costs

The main upgrades of the plant for BNR occurred in 1979, when the Advanced Wastewater Treatment (AWT) plant was installed for phosphorus removal, and in 1997, when the aeration basins were retrofitted for step-feed operation to accomplish nitrogen removal. The AWT is a chemical phosphorus-removal facility that includes mixing and reaction tanks with filtration. The step-feed retrofit consisted of piping modifications and tank additions and filtration.

The costs for installation of the AWT facility were not available; however, they would have been typical of retrofits where chemical is added before tertiary clarifiers and filters because such facilities would be used for normal BOD/TSS removal. This means that the capital expenditure for a retrofit for chemical phosphorus removal is fairly low because all that would be needed would be storage tanks, pumps, and controls, with many of those possibly available by reusing existing equipment.

Plant personnel provided the estimate that the capital expenditure in 1997 that could be attributed to nitrogen removal is $52.5 million. This estimate was updated to 2007 dollars using the *Engineering News-Record* Construction Cost Index (ENR CCI). The ENR CCI is compiled by McGraw-Hill and provides a means of updating historical costs to account for inflation, thereby allowing comparison of costs on an equal basis. From a Web site provided by the U.S. Department of Agriculture, the ENR index for 1997 was 5,826, while the ENR index for May 2007 was 7,942 (USDA 2007). Multiplying the above results by the ratio 7,942/5,826 obtained the result of $71.6 million in 2007 dollars.

This result was annualized using the interest rate formula for determining a set of annual payments for a present value, given an interest rate and payback period. For this and all other case studies for this report, a 6 percent interest rate and 20-year payback were assumed, resulting in a multiplication factor of 0.0872. The annualized capital cost for nitrogen removal was $6.2 million. This annualized capital for nitrogen removal was used for later unit cost estimates for TN and ammonia nitrogen.

The total capital attributed to BNR in 1997 dollars was $52.5 million, which was adjusted to $71.6 million in 2007 dollars using the ENR index. For this 67-MGD facility, this means the capital expenditure per gallon of BNR treatment capacity is $1.07.

Operation and Maintenance Costs

In all case studies prepared for this document, the O&M costs considered were for electricity, chemicals, and sludge disposal. Labor costs for operation and maintenance were specifically excluded for three reasons:

1. Labor costs are highly sensitive to local conditions, such as the prevailing wage rate, the relative strength of the local economy, the presence of unions, and other factors; thus, they would only confound comparison of the inherent cost of various technologies.
2. For most processes, the incremental extra labor involved in carrying out nutrient removal is recognized but not significant in view of the automatic controls and SCADA system that accompany most upgrades.
3. Most facilities were unable to break down which extra personnel were employed because of nutrient removal and related overtime costs, making labor cost development difficult.

The Noman M. Cole, Jr., plant uses primarily chemical phosphorus removal and biological nitrogen removal. This means that the primary O&M costs for phosphorus removal are for electricity, chemicals, and sludge disposal, while the primary O&M costs for nitrogen removal are for electricity. Chemical sludge is recycled to the plant headworks, but it contributes to the eventual primary sludge.

The Attachment lays out the electrical usage for the plant. The entire electrical usage for phosphorus removal lies in the AWT portion of the plant, at 280,000 kilowatt-hours (kWh) per month, or 3,360,000 kWh/yr. Using the average electrical rate of $0.055/kWh, which includes all demand charges, the cost of electricity for phosphorus removal is $185,000. The power usage for nitrogen removal was 18,059,000 kWh/hr. At the average electrical rate, the cost of electricity for nitrogen removal is $993,300.

Plant personnel estimated that chemical (ferric chloride) usage for phosphorus removal cost $1,076/day. In addition, plant personnel estimated that the ferric chloride generated an additional 2 dry tons of primary sludge per day, which cost an additional $1,076/day for disposal. This meant that the additional cost for phosphorus removal for chemical and sludge disposal totaled $785,500/yr. Over the evaluation period, plant personnel used an estimated $250,000 worth of caustic for pH adjustment, which is needed for nitrogen removal.

Unit Costs for Nitrogen and Phosphorus Removal

During the evaluation period, the plant removed 909,600 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal was $1.07/lb, while the annualized unit capital cost for phosphorus removal was $0.

During the evaluation period, the plant removed 4,240,000 lb of TN. With the results above, the unit O&M cost for TN removal was $0.29/lb of TN, while the annualized unit capital cost for TN removal was $1.47.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle cost is the sum of the annualized unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal was $1.07/lb and the life-cycle cost for TN removal was $1.76/lb.

Assessment of magnitude of costs: The capital cost of $1.07/gpd capacity is low because of the existing facility before the upgrade. The O&M cost for phosphorus removal is high due to chemical use to reach a low concentration limit, while the O&M cost for nitrogen removal are in the middle range, compared with those for other facilities.

Summary

The Noman M. Cole, Jr., plant retrofit to a step-feed strategy has provided excellent reliability in meeting both nitrogen and phosphorus limits. The COVs were 21 percent for

TP at the annual average of 0.09 mg/L, 14 percent for ammonia nitrogen at the average concentration of 0.12 mg/L, and 12 percent for TN at the average concentration of 5.25 mg/L. The phosphorus removal is achieved primarily by chemical addition followed by tertiary filters. The nitrogen removal is achieved with multiple anoxic zones in the process. In addition, the step-feed provides operational benefits during wet-weather conditions because the strategy allows the operators to distribute the increased flows throughout the aeration basins in steps, thereby protecting the clarifiers from added solids loadings during high-flow periods. Removal costs for both phosphorus and nitrogen were reasonable, with low capital at $1.07/gpd capacity, and O&M costs at $1.07/lb TP removed and $1.77/lb TN removed.

Acknowledgments

The authors are grateful for the significant help and guidance provided by Michael McGrath, operations director, and Roger Silverio, process engineer, at the Noman M. Cole, Jr., Pollution Control Plant. This case study would not have been possible without their prompt response with well-deserved pride in the facility and its operation. The authors also acknowledge Fairfax County for its participation in this case study.

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Attachment: Facility Design Information

Design Flow	
Minimum flow	26.8 MGD
Average daily flow	67.0 MGD
Peak instantaneous flow	134.0 MGD
Peak process flow	107.2 MGD
Design Average Loadings	
BOD	118,000 lb/d
TSS	126,000 lb/d
TKN	21,000 lb/d
TP	4,100 lb/d
Retention Basin 1 (QQ1)	
Retention QQ1	
Quantity	1
Type	Open
Volume	5.7 MG
Retention basin pumps	
Quantity	4
Type	Submersible
Capacity	
Large	3,300 gpm at 27 ft
Small	350 gpm at 27 ft
Screen Building (B1)	
Bar screens	
Quantity	3
Total channel width	8 ft
Opening size	3/4 in
RAW Wastewater Pump Station (B)	
RAW wastewater pumps	
Quantity	5
Type	Vertical, centrifugal
Speed	
A-1	Adjustable
A-2	Two-speed
A-3	Constant
A-4	Adjustable
A-5	Constant

Capacity	
A-1	20,500 gpm at 30 ft TDH
A-2	19,165 gpm at 30 ft TDH
A-3	20,700 gpm at 30 ft TDH
A-4	20,700 gpm at 30 ft TDH
A-5	18,500 gpm at 30 ft TDH
RAW Wastewater/EQ Tank Pump Station (B2)	
Equalization tank pumps	
Quantity	3
Type	Submersible
Capacity, each	6,544 gpm at 84 ft TDH
Raw wastewater pumps	
Quantity	2
Type	Submersible
Capacity, each	9,682 gpm at 47 ft TDH
Equalization Tanks (B3)	
Equalization tanks	
Quantity	4
Type	Concrete
Dimensions, each	200 ft long X 100 ft wide X 27 ft Deep (SWD)
Volume, each	4 MG
Flash Mix Tanks (C1)	
Quantity	2
Dimensions	30 ft L X 18 ft W X 10 ft SWD
Volume, each	40,400 gallons
Detention time	1.74 minutes at average daily flow
Primary Settling Tanks (C)	
Primary settling tanks	
Quantity	8
Type	Rectangular
Size	139 ft L X 45 ft W X 10 SWD
Weir length, each	120 ft
Weir loading	69,800 gpd/linear foot at average daily flow
Hydraulic overflow rate	1,340 gpd/ft^2 at average daily flow
Primary influent odor control scrubber	
Quantity	1
Type	Packed bed
Depth of packing	12 ft min
Cross section area	19.6 ft^2
Capacity	5,000 CFM

Scrubber recirculation pump	
Quantity	1
Type	Vertical wet pit centrifugal
Capacity, each	100 gpm at 30 ft TDH
Horsepower	2
Small Activated Sludge Tanks 1 TO 6 (D)	
Small activated sludge tanks	
Quantity	6
Number of passes, each	3
Size, each pass	182 ft L X 30 ft W X 13.6 ft SWD
Volume, each tank	1.67 MG
Total volume	10.0 MG
Total anoxic volume	3.5 to 5.0 MG
HRT @ average flow	8.9 hours
SRT @ max mo load, & last pass MLSS OF 4,400	18 days
Mixers	
Quantity	78
Type	Submersible, mast-mounted
Horsepower, each	4 HP
Process oxygen requirements	
BNR operation	
Average	48,000 lb/d
Maximum month	51,000 lb/d
Maximum day	71,400 lb/d
Nitrification only operation	
Average	70,800 lb/d
Maximum month	76,200 lb/d
Maximum day	115,000 lb/d
Diffused aeration equipment	
Type	9-in porous flexible membrane Full floor coverage
Large Activated Sludge Tanks 7 TO 9 (D1)	
Large activated sludge tanks	
Quantity	3
Number of passes, each	6

Size, each pass	
	2 at 165 ft L X 18 ft W X 22 ft SWD
	4 at 165 ft L X 36 ft W X 22 ft SWD
Volume, each tank	4.89 MG
Total volume	14.7 MG
Total anoxic volume	5.1 to 7.3 MG
HRT @ average flow	8.9 hours
SRT @ max mo load,	18 days
& last pass MLSS of 4,400	
Mixers	
Quantity	57
Type	Vertical turbine
	Platform-mounted
Horsepower, each	24 at 3 HP
	12 at 5 HP
	9 at 7.5 HP
	12 at 15 HP
PE channel mixers	
Quantity	18
Type	Submersible, mast-mounted
Horsepower, each	2.5 HP
Process oxygen requirements	
BNR operation	
Average	88,200 lb/d
Maximum month	95,700 lb/d
Maximum day	149,000 lb/d
Nitrification only operation	
Average	101,000 lb/d
Maximum month	108,000 lb/d
Maximum day	164,000 lb/d
Diffused aeration equipment	
Type	9-in porous flexible membrane
	Full floor coverage
AST dewatering pumps	
Quantity	
Large	2
Small	2

Type	Submersible
Capacity, each	
Large	2,025 gpm at 25 ft TDH
Small	75 gpm at 60 ft TDH
Horsepower, each	
Large	25 HP
Small	5 HP
Blower Building (E1)	
Small AST aeration blowers	
Quantity	4
Type	Multistage centrifugal
Capacity, each	16,000 SCFM at 8.0 psi
Horsepower, each	800 HP
Clarifiers 12–15 RAS pumps	
Quantity	5
Type	Single-passage screw impeller, centrifuge
Speed	Adjustable
Capacity, each	4,400 gpm at 28 ft TDH
Horsepower, each	50 HP
WAS pumps	
Quantity	4
Type	Horizontal centrifugal
Capacity, each	510 gpm at 30 ft TDH
Blower Building (E2)	
Aeration blowers	
Quantity	
Small AST blowers	2
Large AST blowers	4
Type	Multistage centrifugal
Capacity	
Small AST blowers	17,500 at 8.0 psi
Large AST blowers	14,000 at 12.7 psi
Horsepower, each	
Small AST blowers	800 HP
Large AST blowers	1,250 HP

Clarifiers 5–8 RAS pumps	
Quantity	5
Type	Horizontal centrifugal
Capacity	6,500 gpm at 37 ft TDH
Clarifiers 16–17 RAS pumps	
Quantity	2
Type	Single-passage screw impeller, centrifuge
Speed	Adjustable
Capacity, each	4,400 gpm at 28 ft TDH
Horsepower, each	50 HP
Secondary Clarifiers 5 to 8 (F)	
Quantity	4
Type	Circular
Diameter	145 ft
Sidewater depth	14.75 ft
Hydraulic overflow rate	540 gpd/ft^2 at peak process flow
Solids loading rate	31 lp/d/ft^2 at peak process flow
Secondary Clarifiers 12 to 17 (F1)	
Secondary clarifiers	
Quantity	6
Type	Rectangular chain & flight
Dimensions, each	260 ft L X 55 ft W X 16 ft SWD
Hydraulic overflow rate	540 gps/ft^2 at peak process flow
Solids loading rate	31 lf/d/ft^2 at peak process flow
Secondary clarifier dewatering pumps	
Quantity	2
Type	Submersible
Capacity, each	500 gpm at 50 ft TDH
Horsepower, each	15 HP
Chlorination Facility (G)	
SPH pumps	
Quantity	4
Type	Vertical turbine
Capacity, each	3,100 gpm at 216 ft TDH
SPH strainers	
Quantity	3
Type	Automatic, self-cleaning
Capacity, each	1,050 gpm

Sodium hypochlorite feed pumps	
Quantity	
Large NaOCl pumps	4
Small NaOCl pumps	2
Type	Tubular diaphragm chemical metering
Capacity	
Large NaOCl pumps	200 gph max
Small NaOCl pumps	50 gph max
Sodium hypochlorite storage tanks	
Quantity	4
Dimensions, each	11.5 ft dia X 15.5 ft high
Volume, each	12,000 gallons
Chemical Feed Building (S)	
Caustic feed pumps	
Quantity	4
Type	Tubular diaphragm chemical metering
Control	Adjustable stroke & speed
Capacity, each	420 gph max
Typical dose	11 mg/L as $CACO_3$ for PH control
Caustic storage tanks	
Quantity	3
Dimensions, each	12 ft diameter X 19 ft high
Volume, each	16,000 gallons
Polymer feed pumps	
Quantity	12
Type	Progressing cavity
Speed	Adjustable
Capacity, each	250 gph max
Typical dose	0.5–1.0 mg/L
Polymer transfer pump	
Quantity	1
Type	Progressing cavity
Capacity	80 gpm

Polymer mixing, aging, and storage tanks	
Quantity	2
Dimensions, each	7 ft dia X 7 ft high
Volume	2,000 gallons
Chemical feed pumps for primary settling tank odor control	
Quantity	
Caustic	1
Sodium hypochlorite	1
Type	Eccentric lobe peristaltic
Capacity	
Caustic	8.6 gpm
Sodium hypochlorite	7.0 gmp
Sodium hypochlorite storage tank (exist)	
Quantity	1
Dimensions, each	12 ft dia X 19 ft high
Volume	16,000 gallons
Equalization Basins 2 & 3 (QQ2 & QQ3)	
Equalization basins	
Type	Concrete-lined, open
Volume	
Basin QQ2	7.4 MG
Basin QQ3	5.8 MG
Wash water return pumps	
Quantity	2
Type	Submersible
Speed	Constant
Capacity, each	600 gpm at 50 ft TDH
Horsepower, each	20 HP
ASE Pump Station (BB)	
ASE pumps	
Quantity	5
Type	Vertical turbine
Capacity	
Adj speed	2 at 29,400 gpm
Constant speed	1 @ 22,600 gpm
Constant speed	2 @ 16,000 gpm

Tertiary Clarifiers (CC)	
Tertiary clarifiers	
Quantity	4
Type	Octagonal
Nominal inside diameter	148 ft
Hydraulic overflow rate	735 gpd/ft^2 at average flow
Tertiary clarifier dewatering pumps	
Quantity	2
Type	Horizontal centrifugal
Speed	Constant
Capacity, each	2,400 gpm at 50 ft TDH
Horsepower	50 HP
Tertiary Clarifiers (CC1)	
Flow distribution structure mixer	
Quantity	1
Type	Vertical turbine, platform-mounted
Horsepower	15 HP
Tertiary clarifier	
Quantity	1
Type	Circular
Diameter	152 ft
Hydraulic loading rate	735 gpd/ft^2 at average flow
Tertiary clarifier dewatering pumps	
Quantity	1
Type	Horizontal centrifugal
Speed	Constant
Capacity, each	1,000 gpm at 21 ft TDH
Horsepower, each	15 HP
TCE Pump Station (CC)	
Tertiary clarifier effluent pumps	
Quantity	3
Type	Vertical turbine
Speed	Adjustable
Capacity, each	22,700 gpm at 35 ft TDH
Horsepower, each	300 HP
Foreign Sludge Incinerator Building (KK)	
Ferric chloride pumps	
Quantity	4
Type	Tubular diaphragm chemical metering

Control	Adjustable stroke & speed
Capacity	200 gpm max
Typical dose	25–30 mg/L
Polymer feed pumps	
Quantity	6
Type	Progressing cavity
Speed	Adjustable
Capacity	2.0 gpm max
Typical dose	0.1–0.2 mg/L
Monomedia Filter Building (FF)	
Monomedia filters	
Quantity	8
Type	Center gullet
Media type	Anthracite
Number cells, each	2
Dimensions, each cell	30 ft L X 17 ft W
Media depth	5 ft
Design loading rate	2.9 gpm/ft^2 at average daily flow with all units in service
Backwash pump	
Quantity	1
Type	Vertical turbine
Capacity	20,400 gpm
Gravity Filter Building (DD)	
Gravity filters	
Quantity	10
Media type	Anthracite/sand
Dimensions, each cell	30 ft L X 30 ft W
Media depth	2.25 ft
Design loading rate	2.6 gpm/ft^2 at average daily flow with all units in service
Backwash pump	
Quantity	1
Type	Vertical turbine
Capacity	18,000 gpm
Gravity filter effluent pumps	
Quantity	
Constant speed	2
Adj speed	2
Type	Vertical turbine

Capacity, each	
Constant speed	22,500 gpm
Adj speed	27,000 gpm
Backwash Effluent Tanks (EE)	
Quantity	3
Dimensions, each	85 ft L X 20 ft W X 11.3 ft SWD
Volume, each	144,000 gallons
Reaeration Tank (HH)	
Quantity	1
Dimensions	72 ft L X 70 ft W X 22 ft SWD
Volume	830,000 gallons
APW Pump Station (HH1)	
Advanced plant water pumps	
Quantity	4
Type	Vertical turbine
Speed	Adjustable
Capacity, each	4,400 gpm at 212 ft TDH
Horsepower, each	300 HP
Blended Sludge Storage Tanks (R1/R2)	
Odor control scrubber system	
Quantity	1
Type	Two-stage, packed-bed wet type
Chemicals treated	NH_3, H_2S
Capacity, each	5,000 cfm
Depth of bedding	7 ft
Cross-sectional area	19.6 ft^2
Chemical feed pumps for odor control	
Quantity	
Caustic	2
Sodium hypochlorite	2
Sulfuric acid	2
Type	Tubular diaphragm, chemical metering
Capacity, each	23 gph
Chemical storage tanks	
Chemical	NAOH
Quantity	1
Dimensions, each	6 ft dia X 10 ft high
Volume, each	2,100 gallons

Chemical	NAOCL
Quantity	1
Dimensions, each	4 ft dia X 11 ft 7 in high
Volume, each	1,000 gallons
Chemical	H_2S0_4
Quantity	1
Dimensions, each	38 in dia X 82 in long
Volume, each	400 gallons
Degritting Building (H1)	
Cyclone separators	
Quantity	6
Capacity, each	465 gpm at 12 psi
Grit classifiers	
Quantity	3
Capacity, each	108 ft^3/hr
Primary Sludge Thickeners (J1/J2)	
Gravity thickeners	
Quantity	4
Type	Circular
Diameter	50 ft
Sidewater depth	10 ft
Flotation Thickeners (Q1/Q2)	
DAF thickeners	
Quantity	3
Type	Rectangular
Size	40.2 ft L x 12 ft W x 12 ft SWD
Capacity, each	960 gpm
Sludge Storage (R1/R2)	
Sludge storage tanks	
Quantity	2
Diameter	
Sidewater depth	
Volume, each	367,000 gallons
Sludge Dewatering (K3)	
Centrifuge	
Quantity	4
Type	Bowl and scroll conveyor
Sludge loading, each	
With lime	5,351 lb/hr
Excluding lime	4,730 lb/hr
Sludge feed concentration (percent)	

Minimum	3.00%
Maximum	6.00%
Minimum cake solids concentration	29.00%
Minimum solids capture	95.00%
Capacity, each, based on 3.5% solid feed	
95% solids capture, 29% cake solid	60 dry tons per day
Sludge Incineration (K1/K2)	
Incinerators Nos. 1 & 2	
Quantity	2
Type	Multiple hearth
Capacity, each	45 dry tons per day
Incinerators Nos. 3 & 4	
Quantity	2
Type	Multiple hearth
Capacity, each	92 dry tons per day

North Cary Water Reclamation Facility
North Cary, North Carolina
Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

The North Cary, North Carolina, Water Reclamation Facility (WRF) is a 12-million-gallon-per-day (MGD) facility that included biological nutrient removal (BNR) as part of a 1997 expansion. This facility, which was a replacement/expansion of the 4-MGD Schreiber process on the same site, was selected as a case study because of its phased isolation ditch (PID) technology with tertiary filters.

The WRF does not have primary settling and uses the PID technology or the BioDenipho process by Kruger. The facility uses two pairs of oxidation ditches with anaerobic selectors ahead of the ditches and a second anoxic zone following the ditches. Each pair of ditches is operated in an aerobic/anoxic sequencing mode or phases. The effluent from the ditches goes to two 130-foot-diameter clarifiers. Before discharge to Crabtree Creek, effluent is passed through an upflow Dynasand filter by Parkson and ultraviolet disinfection and is aerated. The original Schreiber tank was converted into a 7-million-gallon (MG) equalization basin in addition to a 2-MG equalization basin at the headworks area, and the stored water is drained by gravity to the influent pump station for subsequent treatment. Sludge is thickened and aerobically digested before it is transported to the South Cary WRF for dewatering and drying for final disposal.

The relevant permit limits that the North Carolina Department of Environment and Natural Resources (NCDENR) established for the plant are shown in Table 1. Compliance limits are primarily for the monthly averages shown for carbonaceous biochemical oxygen demand (CBOD), total suspended solids (TSS), and ammonia nitrogen. Additional limits are specified for the quarterly limit for total phosphorus (TP) and for the annual maximum limit of 144,000 lb for total nitrogen (TN), which is equivalent to 3.94 milligrams per liter (mg/L) as nitrogen.

A distinguishing feature of the BioDenipho process is the alternating flow pattern and process conditions (aerobic and anoxic) occurring within the oxidation ditches. This operating strategy allows nitrogen and CBOD removal to occur within the active process volume, eliminating the need for internal recycle pumping. The operation is executed by a programmable logic controller (PLC)-based system that coordinates the operation of the mechanical process equipment and controls the phase lengths within each ditch. The PLC system allows both manual and automatic control of the treatment process. The PLC panel

Table 1. NCDENR permit limits

Parameter	Summer limits (mg/L)	Winter limits (mg/L)	Quarterly limits (mg/L)	Annual limits
CBOD	4.1	8.2	--	--
TSS	30	30	--	--
NH_3-N	0.5	1.0	--	--
TN	--	--	--	144,000 lb (max)[a]
TP	--	--	2.0	--
Coliforms	--	--	200/100 mL	--

Notes:
NH3-N = ammonia nitrogen
[a] Equivalent to 3.94 mg/L as TN for 12 MGD

also includes preprogrammed operational modes, such as the stormwater mode to address infiltration/inflow (I/I) concerns. For example, automatic or manual activation of the stormwater mode incorporates a sedimentation phase into the BioDenipho process to prevent solids washout during severe rain events. This innovation allows reduction of the required size of the secondary clarifiers or eliminates the requirement for redundant clarifiers.

Plant Design and Process Parameters

A schematic for the North Cary WRF is shown in Figure 1. To ensure economical and efficient treatment, the system also controls the aeration equipment by automatic dissolved oxygen (DO) control. DO probes continuously monitor and report residual DO levels within the oxidation ditches to the PLC panel that controls the aeration equipment to meet, but not exceed, the current oxygen demand. This eliminates costly and wasteful over-aeration that can compromise process stability and operational budgets. Table 2 and Attachment 1 present relevant design data for the facility and Attachment 2 presents a plant operating process diagram. The sludge residence time (SRT) for an oxidation ditch was 12 days at 12 degrees Celsius (°C).

Table 2. Facility design data

Units	Number	Volume
Anaerobic selectors	4 each train	0.093 MG x 4 = 0.372 MG
Oxidation ditch	2 each train	1.5 MG x 2 = 3 MG
Secondary anoxic zone	3 each train	0.111 MG x 3 = 0.333 MG
Reaeration zone	1 each train	0.111 MG
Clarifiers	2 each	130 ft diameter

Note: MG = million gallons

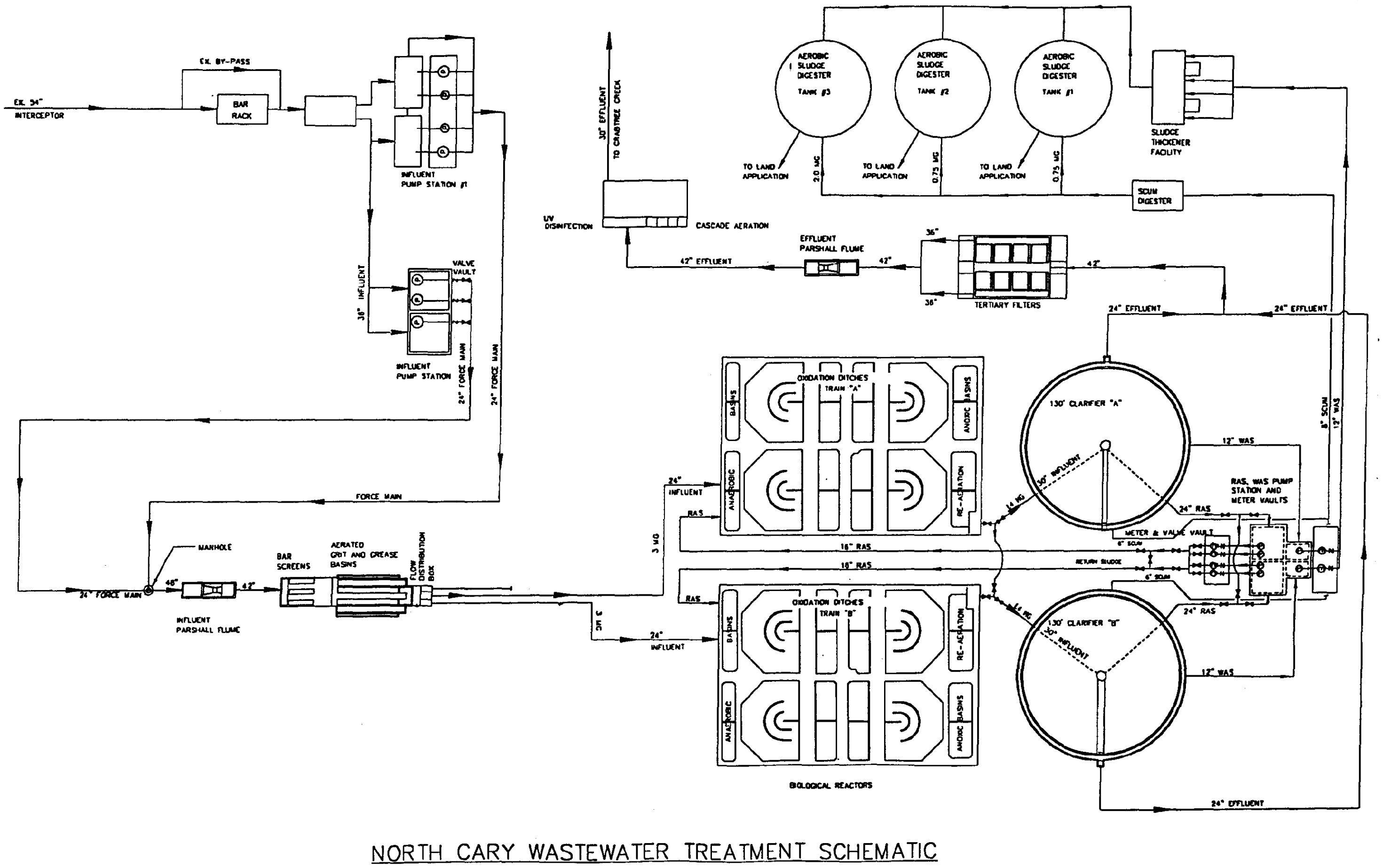

Figure 1. North Cary WRF process schematic.

Table 3 presents operational results for the October 2005 to September 2006 period. Table 4 presents plant monthly average plant process parameters.

Table 3. Influent and effluent averages

Parameter (mg/L unless stated)	Average value	Maximum month	Max month vs. avg.	Maximum week	Sample method/frequency
Flow (MGD)	7.0	8.71	24%	10.8	--
Influent TP (mg/L)	7.7	9.2	20%	11.1	Composite, 3x/week
Effluent TP (mg/L)	0.38	1.06	180%	1.45	Composite, 3x/week
Influent BOD (mg/L)	244	271	11%	296	Composite, 5x/week
Effluent BOD (mg/L)	0.8	1.26	50%	1.84	Composite, 5x/week
Influent TSS (mg/L)	366	418	14%	594	Composite, 5x/week
Effluent TSS (mg/L)	1.0	1.47	45%	2.28	Composite, 5x/week
Influent NH_4-N (mg/L)	45.5	49.4	8%	53.5	Composite, 5x/week
Effluent NH_4-N (mg/L)	0.08	0.34	316%	1.03	Composite, 5x/week
Influent TKN (mg/L)	56.4	62.2	10%	65.6	Composite, 3x/week
Effluent TN (mg/L)	3.67	4.46	21%	5.87	Composite, 3x/week

Note:
TKN = total Kjeldahl nitrogen
BOD = biochemical oxygen demand

Table 4. Monthly averages for plant process parameters

Month	MLSS (mg/L)	Sludge age (days)	HRT (hours)	Temperature (°C)
Oct 2005	2,665	13.1	28	23
Nov 2005	2,628	13.8	29	20
Dec 2005	2,736	13.0	26	19
Jan 2006	2,672	13.3	27	18
Feb 2006	2,720	12.8	27	16
Mar 2006	2,692	13.3	29	18
Apr 2006	2,661	12.6	27	19
May 2006	2,625	13.5	28	21
June 2006	2,700	11.3	21	24
July 2006	2,713	12.3	25	26
Aug 2006	2,709	12.6	25	27
Sep 2006	2,685	12.1	24	26

Notes:
HRT = hydraulic retention time
MLSS = mixed liquor suspended solids

Plant Performance

This section of the case study provides information about the operational performance of nutrient removal at the facility. Figures 2 and 3 present monthly and weekly reliability data for ammonia nitrogen removal. These data cover the period of October 2005 through September 2006. Note that the apparent outlier values are from the period in June 2006 when the plant's service area was subjected to nearly 8 inches of rain in a 24-hour period from Tropical Storm Alberto. Note also that despite that upset, the plant still met the monthly limit of 0.5 mg/L for ammonia nitrogen. Overall, ammonia nitrogen oxidation was complete, with a mean of 0.06 mg/L and a 31 percent coefficient of variation (COV) for non-tropical storm months.

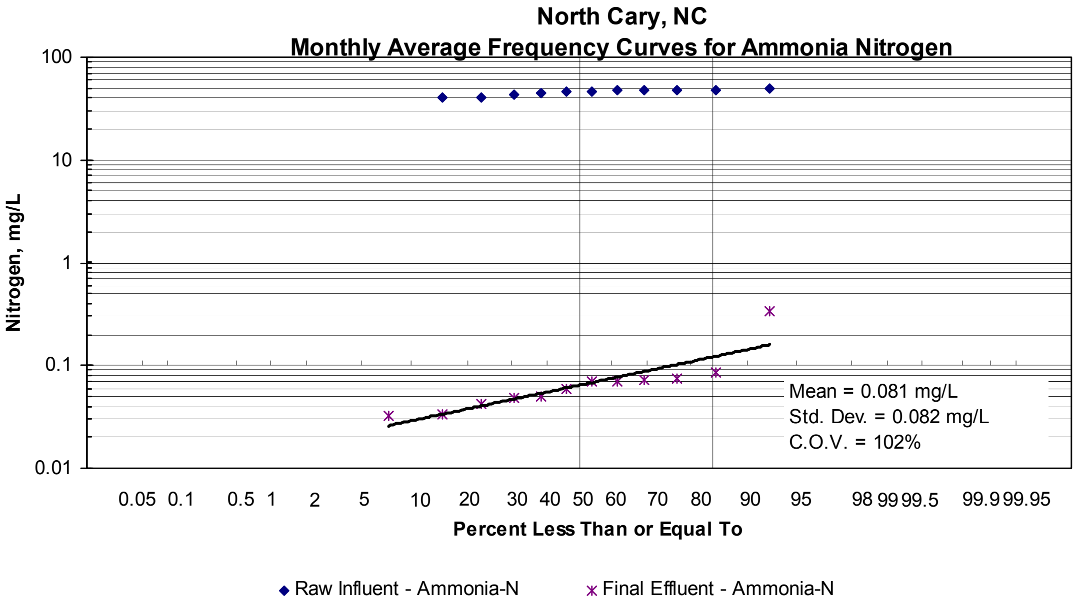

Figure 2. Monthly average frequency curves for ammonia nitrogen.

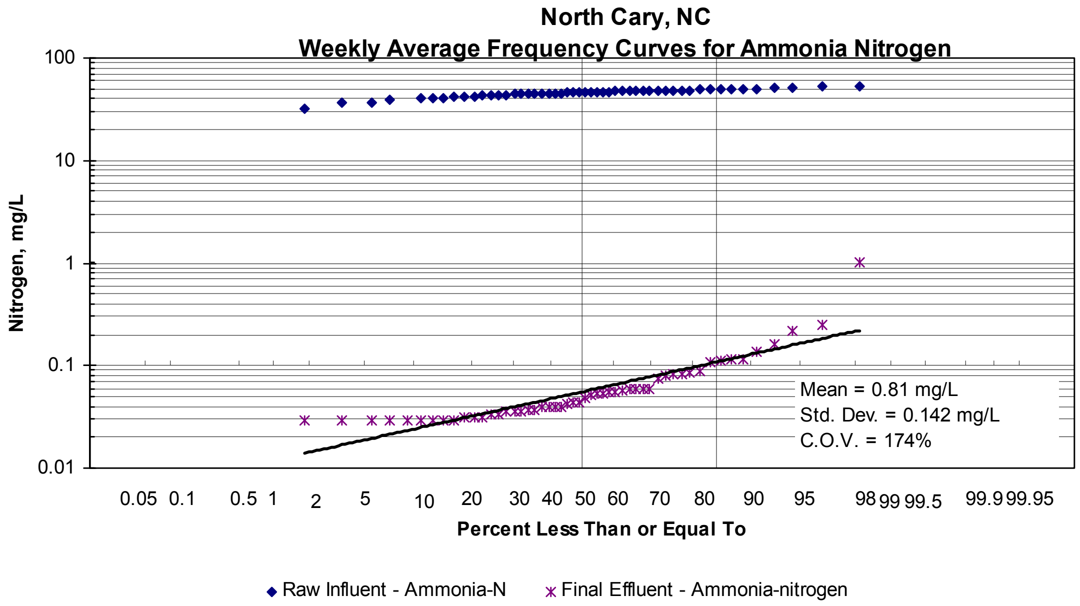

Figure 3. Weekly average frequency curves for ammonia nitrogen.

Figures 4 and 5 present monthly and weekly reliability data for TP removal. Phosphorus removal was completely by biological means and worked well, with a monthly mean of 0.38 mg/L and a COV of 64 percent. This removal was sufficient to meet the facility's quarterly limit of 2 parts per million (ppm).

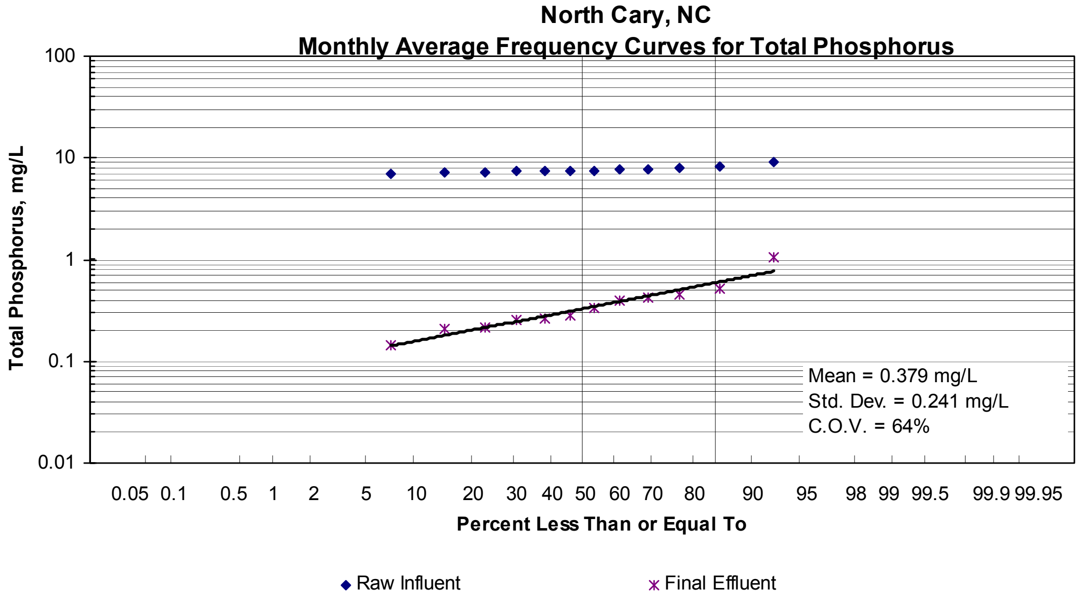

Figure 4. Monthly average frequency curves for TP.

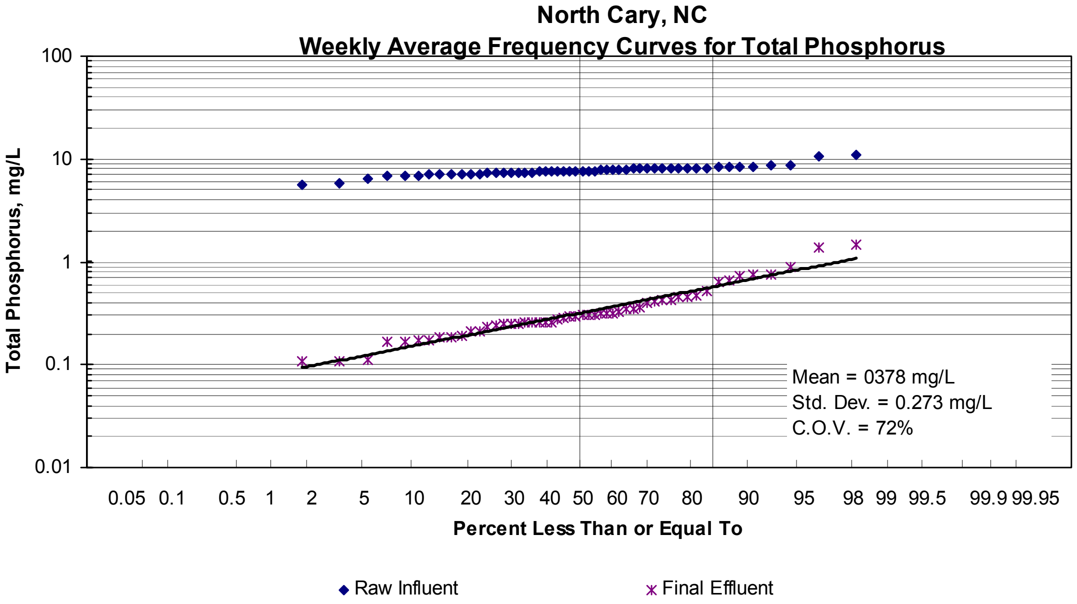

Figure 5. Weekly average frequency curves for TP.

Figures 6 and 7 present reliability data for removal of TN at the facility. TN removal was excellent, with the effluent mean 3.7 mg/L with a COV of 14 percent on a monthly average basis, including the period with heavy precipitation caused by the tropical storm.

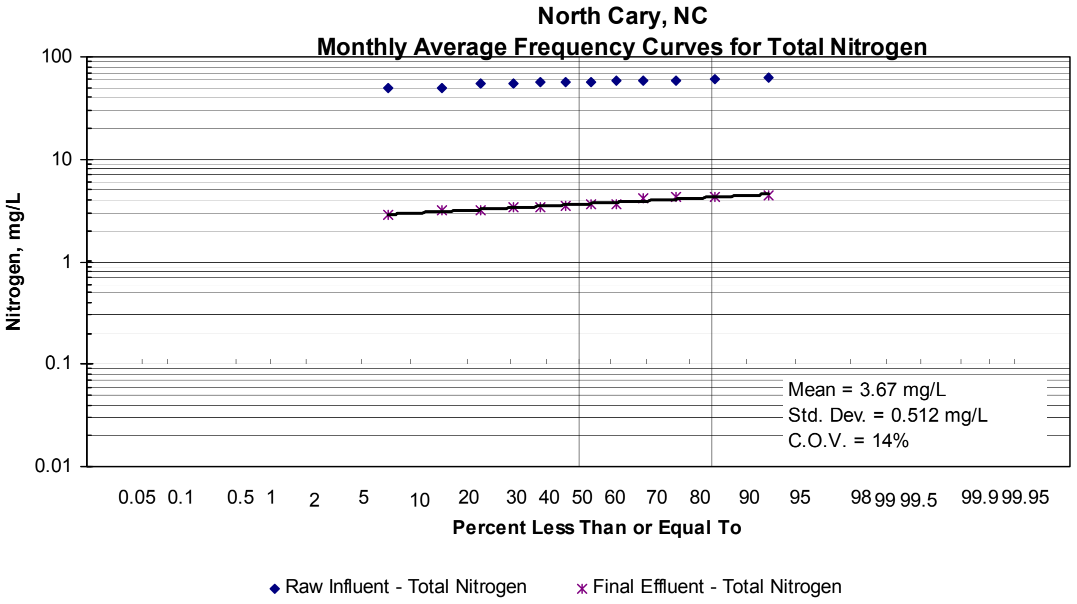

Figure 6. Monthly average frequency curves for TN.

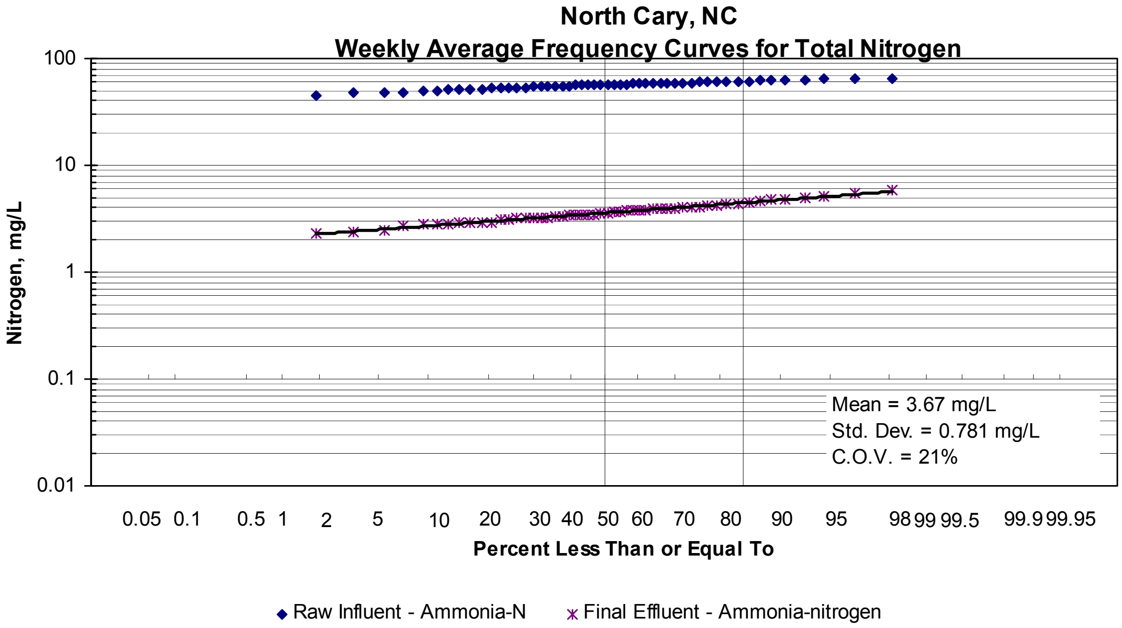

Figure 7. Weekly average frequency curves for TN.

Reliability Factors

The performance was efficient and reliable for entirely biological phosphorus and nitrogen removal at North Cary. The COVs were 102 percent for ammonia nitrogen at the mean concentration of 0.08 mg/L, 64 percent for total phosphorus at the mean concentration of 0.38 mg/L, and 14 percent for total nitrogen at the mean concentration of 3.67 mg/L.

The following points summarize the factors affecting the reliability of the North Cary WRF:

- The BioDenipho process at North Cary is a flexible process with regard to varying wastewater strength and flow rate. The reliability is achieved through well-controlled oxidation of ammonia and subsequent denitrification in two distinct anoxic steps. The anoxic cycle phase in the ditch can be adjusted from 60 minutes to 90 minutes, for example, during a low-flow period, while it can be reversed during a high-flow period. The rotors are controlled to provide sufficient oxygen to maintain the DO concentration at 1 to 1.5 mg/L in the ditch, while mixers keep the organisms in suspension during the anoxic phase. This flexibility to control mixing separately from aeration is one of the keys to this plant's reliability. The low DO in the ditch effluent ensures good denitrification in the second anoxic step to reach the desired nitrogen level in the effluent. No external carbon source is needed to meet the permit limit.

- Another key reliability factor is the automated control system, which consists of sensors and DO controllers operating with the PLC and associated supervisory control and data acquisition (SCADA) system. The exact phasing decision is made on the basis of the preset control logic, which is site-specific and fully automated.
- A key reliability factor for biological phosphorus removal is the feed point of the influent. The influent is fed to the second anaerobic selector, while return activated sludge is fed to the first selector to ensure that the returning nitrate from the clarifier will be denitrified in the first selector zone. The second, third, and fourth selector zones thus become anaerobic and allow full energy exchange for polyphosphate-accumulating organisms (PAOs). The wastewater exhibited a favorable ratio of biochemical oxygen demand (BOD) to TP, greater than 30 as an average. The plant performance has been proven reliable through this process (WEF and ASCE 1998).
- A key reliability factor for nitrogen removal is the three phases of anoxic cycles. The first is in the anaerobic selector before the ditch, the second is in the ditch, and the third is in the anoxic zone after the ditch. These multiple opportunities to denitrify in the presence of BOD in the wastewater are unique and ensure good removal of nitrogen. The wastewater exhibited a favorable BOD/TKN ratio of 5 as an average, which is adequate for good denitrification (USEPA 1993).
- Training is another key factor for achieving high reliability. Online monitoring and automatic controls make training easy but require continuous maintenance by the plant personnel.
- Less power is used because of the maximum use of nitrate during the anoxic phase and the prevention of over-aeration during the oxic phase. Pumping of oxidized effluent to 3 to 4 times the discharge (Q) is not required to reach the same level of denitrification.
- Tertiary filters are effective in suspended solids removal.
- Recycle loads are minimized; aerobic digestion occurs on-site, and the digested sludge is shipped away for processing at another facility.
- Wet-weather flows are handled in two ways: The equalization basins have a total of 9 MG storage, or 75 percent of the influent design flow; the PID has a storm mode in the process control, under which the program switches into a sedimentation phase, thereby preventing solids washout. These helped manage high flows during the June 2006 event, when Tropical Storm Alberto brought high flows to the plant. All the storage volume was used, and the PID went into the storm mode for a short duration. The plant treated all flows and complied with the permit.

Costs

Capital Costs

The main upgrade of the plant for BNR was in 1997 when the ditches were installed. The upgrade then cost $25 million. The upgrade included additional ditches, the selector, pumps, aerators, and tertiary filtration.

Because all phosphorus and nitrogen removal is biological, the capital costs were attributed to different removal processes on the basis of the amount of oxygen used during biological treatment, which is 12 percent for TP removal, 48 percent for nitrogen removal, and 40 percent for other (i.e., BOD removal). This means that the capital expenditure attributed to TP removal was $3 million, and the expenditure attributed to nitrogen removal was $12 million.

These capital cost results were updated to 2007 dollars using the *Engineering News-Record* Construction Cost Index (ENR CCI). The ENR CCI is compiled by McGraw-Hill and provides a means of updating historical costs to account for inflation, thereby allowing comparison of costs on an equal basis. From a Web site provided by the U.S. Department of Agriculture, the ENR index for 1997 was 5,826, while the ENR index for May 2007 was 7,942 (USDA 2007). Multiplying the above results by the ratio 7,942/5,826 obtained the result of $4.09 million for phosphorus removal and $16.9 million for nitrogen removal in 2007 dollars.

The total capital expenditure attributed to BNR in 2007 dollars was $34.1 million. For the 12-MGD facility, the capital expenditure per gallon of BNR treatment capacity was $2.84.

Operation and Maintenance Costs

In all case studies prepared for this document, the O&M costs considered were for electricity, chemicals, and sludge disposal. Labor costs for O&M were specifically excluded for three reasons:

1. Labor costs are highly sensitive to local conditions, such as the prevailing wage rate, the relatively strength of the local economy, the presence of unions, and other factors; thus, they would only confound comparison of the inherent cost of various technologies.
2. For most processes, the incremental extra labor involved in carrying out nutrient removal is recognized but not significant in view of automatic controls and SCADA system that accompany most upgrades.
3. Most facilities were unable to break down which extra personnel were employed because of nutrient removal and related overtime costs, making labor cost development difficult.

The plant uses an entirely biological phosphorus removal process to achieve the limit; therefore, the primary operating cost is electrical use for operating the mixers, pumps, and selector. Attachment 3 presents the electrical cost calculations for one train; the second train is a duplicate. Power usage was attributed on the basis of discussions with plant personnel, who suggested 5 percent for phosphorus removal and 95 percent for nitrogen removal, except for units that could be entirely attributed to phosphorus or nitrogen (i.e., anaerobic mixers for phosphorus, anoxic mixers for nitrogen). From this, the total power usage attributed to phosphorus removal was 377,000 kilowatt-hours per year (kWh/yr). When calculated using the average electrical cost of $0.056/kWh (which includes all demand charges), the cost for power for phosphorus removal was $17,400. The total power usage attributed to nitrogen removal was 2,558,000 kWh/yr; applying the electrical unit price, the cost for power for nitrogen removal was $118,000.

The sludge generated during the process is transported to another town of Cary facility for disposal. From consultation with plant personnel, the sludge generated (4.91 tons/day) was attributed at 5 percent to phosphorus removal and 95 percent to nitrogen removal. The cost for the plant to send the sludge out for treatment was $200/ton. The cost for sludge disposal for phosphorus removal was $17,900, while the sludge disposal for nitrogen removal was $341,000.

Unit Costs for Nitrogen and Phosphorus Removal

During the evaluation period, the plant removed 156,000 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal was $0.23/lb, while the annualized unit capital cost for phosphorus removal was $2.28.

During the evaluation period, the plant removed 1,121,000 lb of total nitrogen. With the results above, the unit O&M cost for total nitrogen removal was $0.41/lb of TN, while the annualized unit capital cost for TN removal was $1.27.

Life-Cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the annualized unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal was $2.51/lb and the life-cycle cost for TN removal was $1.68/lb.

Assessment of magnitude of costs: The capital cost of $2.84 per gallon per day (gpd) capacity is relatively high, but the O&M costs are very low. One of the key factors is that chemicals are not used for nutrient removal, saving both those costs as well as costs that would be attributed to additional sludge generation.

Summary

The North Cary facility is unique in that it provides reliable nutrient removal by means of a PID process followed by tertiary filters. The phosphorus removal is achieved entirely by a biological process with a mean concentration of 0.38 mg/L with a COV of 64 percent. The nitrogen removal is also achieved entirely by a biological process with a mean of 3.67 mg/L with an extremely low COV of 14 percent. The process is flexible enough to accommodate varying flow conditions and the wastewater characteristics through the year, including the severe rain caused by Tropical Storm Alberto in June 2006. Automatic controls incorporated into the plant ensure reliable operation and control through these operating periods. The wastewater characteristics are favorable to both nitrogen and phosphorus removal, and no external carbon sources are needed with this PID process.

The capital cost is relatively high at $2.84/gpd capacity as a new facility but compares well with others, which normally exceed $3/gpd. The O&M costs are estimated at $1.26/lb of TP removed and $0.41/lb of TN removed. These costs are remarkably low, reflecting the inherent advantages of this unique treatment process. The total costs were $2.21/lb of TP removed and $2.92/lb of TN removed.

Acknowledgments

The authors are grateful for the significant assistance and guidance that Chris Parisher, North Cary WRF superintendent, provided. This case study would not have been possible without his prompt response with well-deserved pride in the facility and its operation. The authors also wish to thank the town of Cary for its participation.

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Attachment 1: Key Design Parameters

SECTION 5: KEY DESIGN PARAMETERS

Design Basis:

	Influent	Expected Secondary Effluent	Unit
Annual Average Daily Flow	10	--	MGD
Peak Daily Flow	20	--	MGD
BOD	250	<10	mg/l
TSS	300	<10	mg/l
TKN	35	--	mg/l
NH;N’	--	<0.5/1.0	mg/l
TP	7	2	mg/l
TN	--	6	mg/l
Temp	12/27	--	ºC

***Summer/Winter**

System Parameters:

		Unit
Number of Trains	2	
Anaerobic Selector		
Physical Parameters		
Number of Stages/Train	4	
Volume per Stage	0.093	MG
Length per Stage	35.3	ft
Width per Stage	17.6	ft
SWD	20.0	ft
Equipment/Stage		
Mixers		
Number	1	
Model	POP-I	
HP	4.9	HP
RPM	180	RPM

Attachment 2: Operating Stages of the BioDenipho Process

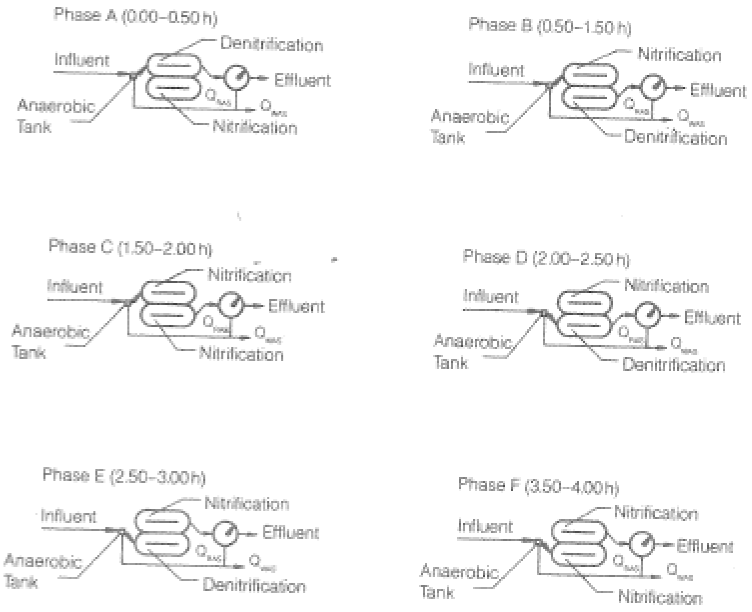

Figure 8. Operating stages for the BioDenipho process (WEF and ASCE 1998).

Attachment 3: Electrical Costs

Electrical	1 train								
Hp	Number	kW Power draw	hours/day	kWh draw/day	kWh draw/year	% for P	% for N	for P	for N
Anaerobic Mixers									
4.9	4	14.6216	24	350.9184	128,085.2	100	0	128,085.2	0
Rotors									
60	4	179.04	15.12	2,707.085	988,086	5	95	49,404.3	938,681.65
Main mixers									
9	4	26.856	8.88	238.4813	87,045.67	5	95	4,352.283	82,693.384
Anoxic mixers									
6.5	3	14.547	24	349.128	127,431.7	0	100	0	127,431.72
Reaeration blower									
20	1	14.92	24	358.08	130,699.2	5	95	6,534.96	124164.24
Clarifer drive									
1	1	0.746	24	17.904	6,534.96	5	95	326.748	6,208.212
Total for 1 train								188,703.5	1,279,179.2
Total for 2 trains	TRAINS							377,407	2,558,358.4
					Rate	0.05		for P	for N
					Totals			$17,361	$117,684

Western Branch Wastewater Treatment Plant Upper Marlboro, Maryland

Nutrient Removal Technology Assessment Case Study

Introduction and Permit Limits

The Western Branch Wastewater Treatment Plant (WWTP) was selected as a case study because of a unique feature—three separate activated-sludge systems operating in series to remove nutrients.

The Western Branch WWTP is part of the Washington Suburban Sanitary Commission (WSSC), and it is in Upper Marlboro, Maryland. It is permitted for a flow of 30 million gallons per day (MGD); in 2006 it processed an average of 19.3 MGD. The plant is permitted to discharge to the Western Branch of the Patuxent River.

The relevant National Pollutant Discharge Elimination System (NPDES) permit limits for the facility are shown in Table 1.

Table 1. NPDES permit limits

Parameter	Annual loading (mg/L)	Monthly average (mg/L)	Weekly average (mg/L)
BOD_5 4/1–10/31		9	14
BOD_5 11/1–3/31		30	45
TSS		30	45
Total phosphorus	0.3	1.0	N/A
Total nitrogen	4.0	3.0	4.5
Ammonia-N 4/1–10/31		1.5	N/A
Ammonia-N 11/1–3/31		5.5	N/A

Notes:
BOD = biochemical oxygen demand
mg/L = milligrams per liter
N/A = not applicable
TSS = total suspended solids
Note that 0.3 mg/L TP and 4 mg/L TN on an annual load basis will be required after completion of enhanced nutrient removal upgrades funded by Maryland.
[a]Total nitrogen and total phosphorus are based on a design flow of 30 MGD.

Plant Process

Figures 1 is an overall process flow diagram, and Figure 2 is a detailed liquid side process flow diagram for the Western Branch Facility. The plant has three separate activated-sludge systems in series: a high-rate activated-sludge (HRAS) system, intended primarily for BOD removal; a nitrification activated-sludge (NAS) system, for conversion of ammonia nitrogen to nitrate; and a denitrification activated-sludge (DNAS) system, for conversion of nitrate to nitrogen gas. The return activated sludge for each system is kept separated to allow for independent setting of sludge residence times. The system does not include primary settling, and grit removal and screenings are provided ahead of the HRAS. The effluent is filtered prior to ultraviolet (UV) disinfection. Waste activated sludge from the three systems is mixed, thickened by dissolved air flotation (DAF), dewatered by centrifuge, and incinerated in two multiple-hearth incinerators. Process water from the DAF, centrifuge, and incinerator air scrubbers is returned to the headworks.

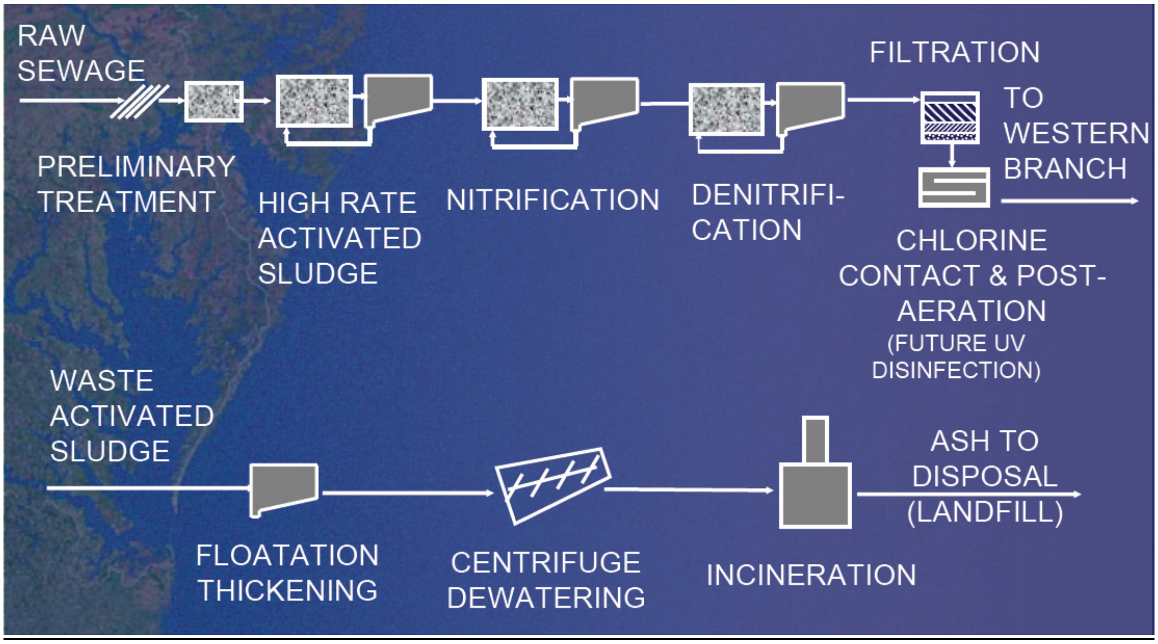

Figure 1. Western Branch WWTP process flow.

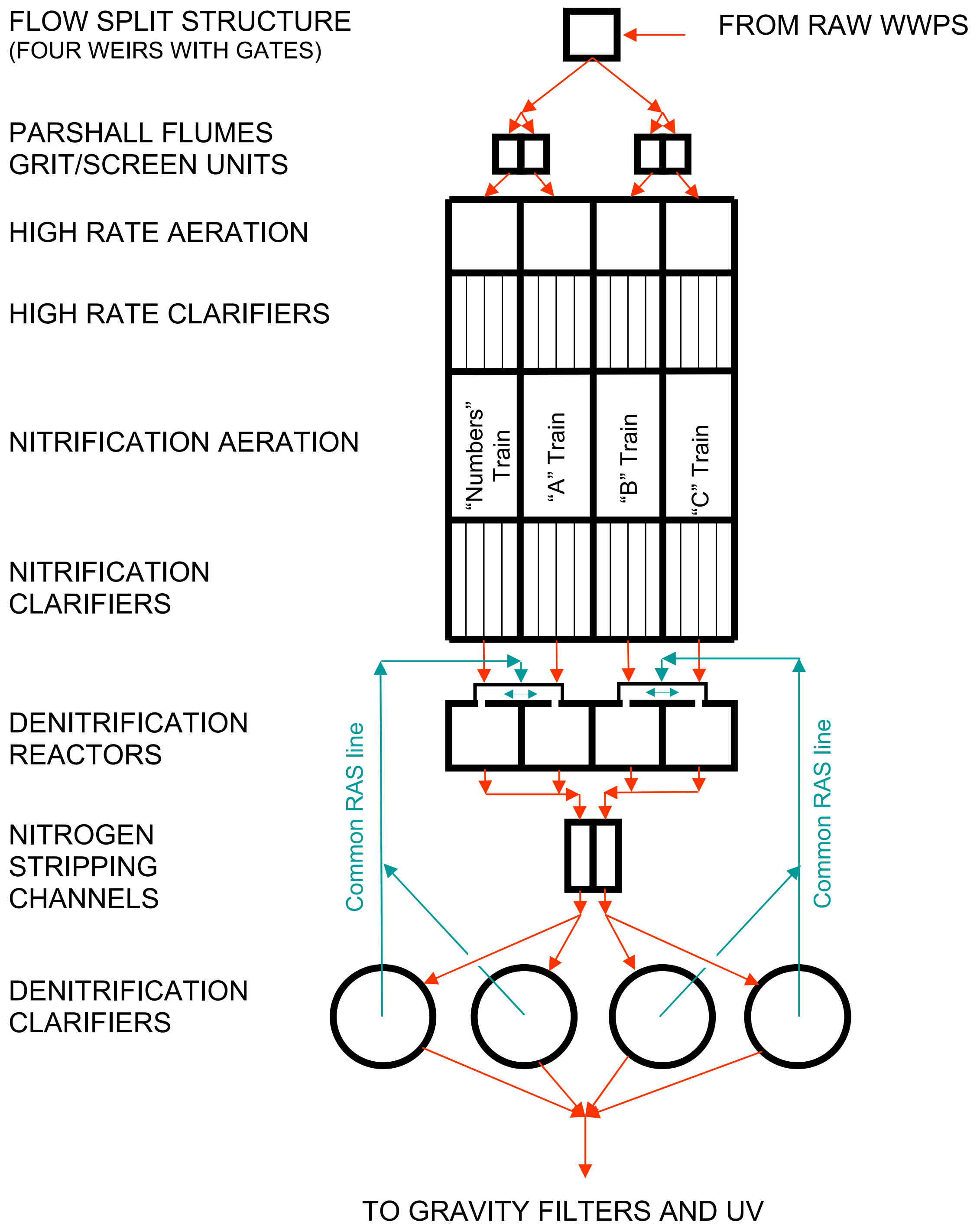

Figure 2. Western Branch WWTP liquid process flow.

Basis of Design and Actual Flow

Flow

The design flow for the facility is 30 MGD. The average flow for the study period was 19.3 MGD (23.0 MGD including recycles), while the maximum month flow during the study period was 26.5 MGD (including recycles) during November 2006.

Loadings

Plant design based on the following:

Plant influent:	BOD	200 mg/L
	TSS	200 mg/L
HRAS effluent:	BOD	60 mg/L
NAS effluent:	BOD	20 mg/L
	TKN	2 mg/l L
	Nitrate-nitrogen	15–30 mg/L

Process	Size	Detention time (hours)	VLR BOD lb/kcf/d
HRAS	3.35 (0.84 MG, 4 each)	2.68	112
NAS	6.89 (1.72 MG, 4 each)	5.51	16
DNAS –Anoxic	3.35 (0.84 MG, 4 each)	2.68	

- Stripping/reaeration = 0.68 MG (0.28 MG, 2 each) 0.45
- TKN loading rate = 5.5 lb TKN/kcf/d
- Sludge age = 5–10 days
- RAS = 100% of plant flow
- Methanol reed rate = 100 mg/L
- Alum feed point is the stripping/reaeration channel

Clarifiers	Size	Overflow rate	SLR
HRAS	120 x 80 x 13 ft, 4 each	781 gpd/ft^2	34 lb/ft^2/d
NAS	150 x 80 x 11.5 ft, 4 each	625 gpd/ft^2	27.4
DNAS	Diameter – 160 ft, 4 each	373	16.3

Note:
SLR = sludge loading rate and is based on a mixed liquor suspended solids concentration of 3,000 mg/L.

Tertiary filters–gravity filters, with air-water backwash capability

- 30 ft x 30 ft, 11 each, total area 9,900 ft^2
- Filter bottom = Leopold clay tiles
- Media–20 inches of anthracite, 8 inches of sand, 12 inches of gravel
- Hydraulic loading rate = 2.1 gpm/ft^2

Plant Parameters

Overall plant influent and effluent average results for the period January 2006 to December 2006 are shown in Table 2.

Table 2. Influent and effluent averages

Parameter (mg/L unless stated)	Average value	Maximum month	Max month vs. ave.	Maximum week	Sample method/frequency
Flow incl recycle (MGD)	23.0	26.5	15%	30.9	
Influent TP	3.70	4.22	14%	5.57	Twice weekly/ composite
Effluent TP	0.43	0.89	89%	0.99	Five times weekly/ composite
Influent COD	332	417	26%	641	Twice weekly/ composite
Effluent COD	16.1	25.8	60%	38.6	Five times weekly/ composite
Effluent BOD	2.69	3.94	46%	6.08	Five times weekly/ composite
Influent TSS	222	282	27%	400	Twice weekly/ composite
Effluent TSS	1.23	2.28	85%	4.60	Five times weekly/ composite
Influent NH4-N	19.6	22.3	14%	25.1	Twice weekly/ composite
Effluent NH_4-N	0.22	0.93	323%	3.41	Five times weekly/ composite
Influent Total N	23.9	28.7	20%	44.8	Twice weekly/ composite
Effluent Total N	1.63	2.46	45%	4.22	Five times weekly/ composite

Notes:
BOD = biochemical oxygen demand
Max month vs. average = (max month – average)/average x 100
NH_4-N = ammonia measured as nitrogen
TN = total nitrogen
TP = total phosphorus
TSS = total suspended solids

Tables 3, 4, 5, and 6 present plant monthly averages for the process parameters, as available.

Table 3. Monthly averages for HRAS process parameters

Month	HRAS MLSS (mg/L)	HRAS sludge age (d)	HRAS HRT (hr)
Jan 2006	4,710	1.9	3.4
Feb 2006	4,232	1.8	3.3
Mar 2006	3,808	1.9	3.6
Apr 2006	3,798	1.7	3.7
May 2006	4,208	2.7	3.8
June 2006	5,454	7.3	3.5
July 2006	4,028	1.8	3.5
Aug 2006	4,306	1.9	3.8
Sept 2006	5,545	2.7	3.5
Oct 2006	4,066	1.7	3.4
Nov 2006	3,431	0.9	3.0
Dec 2006	4,017	2.0	3.6

Table 4. Monthly averages for NAS process parameters

Month	NAS MLSS (mg/L)	NAS sludge age (d)	NAS HRT (hr)
Jan 2006	4,264	34.8	7.1
Feb 2006	3,800	29.9	6.7
Mar 2006	3,617	46.6	7.4
Apr 2006	2,794	34.7	7.7
May 2006	3,644	24.3	7.7
June 2006	3,706	21.4	7.2
July 2006	3,523	72.5	7.3
Aug 2006	4,286	65.6	7.9
Sept 2006	4,987	84.6	7.1
Oct 2006	4,806	79.7	6.9
Nov 2006	4,212	34.4	6.2
Dec 2006	5,117	43.2	7.3

Table 5. Monthly averages for DNAS process parameters

Month	DNAS MLSS (mg/L)	DNAS sludge age (d)	DNAS HRT (hr)
Jan 2006	5,006	32.6	3.4
Feb 2006	4,329	24.4	3.3
Mar 2006	3,541	17.8	3.6
Apr 2006	3,818	8.8	3.7
May 2006	2,795	5.8	3.8
June 2006	3,427	11.9	3.5
July 2006	4,201	23.3	3.5
Aug 2006	3,192	10.9	3.8
Sept 2006	3,939	58.4	3.5
Oct 2006	3,968	18.9	3.4
Nov 2006	4,081	40.2	3.0
Dec 2006	4,990	17.0	3.6

Table 6. Monthly averages for influent temperature

Month	Temperature (°F)	Temperature (°C)
Jan 2006	58.5	14.7
Feb 2006	56.3	13.5
Mar 2006	57.5	14.2
Apr 2006	61.9	16.6
May 2006	64.4	18.0
June 2006	68.2	20.1
July 2006	72.4	22.4
Aug 2006	73.6	23.1
Sept 2006	71.4	21.9
Oct 2006	67.7	19.8
Nov 2006	63.9	17.7
Dec 2006	61.1	16.2

Performance Data

Figure 3 presents reliability data for the removal of total phosphorus (TP). The removal is good, with the effluent TP averaging 0.43 mg/L, and a coefficient of variation (COV) of 62 percent.

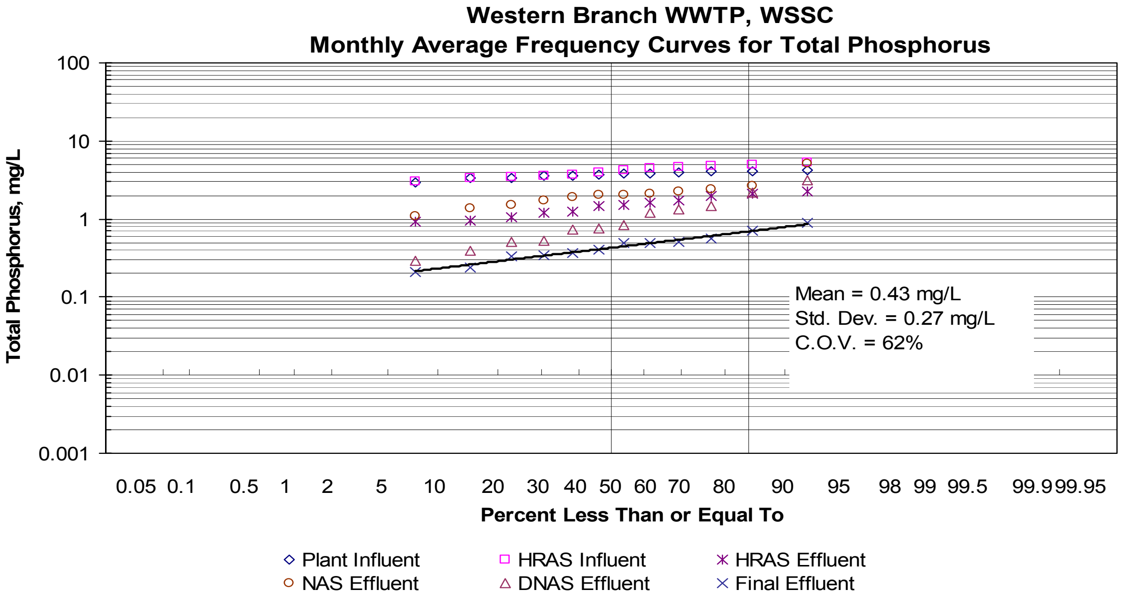

Figure 3. Monthly average frequency curves for TP.

Figure 4 presents reliability data for ammonia nitrogen removal. Removal of ammonia nitrogen is very good, with a mean effluent of 0.13 mg/L and a high COV of 163 percent.

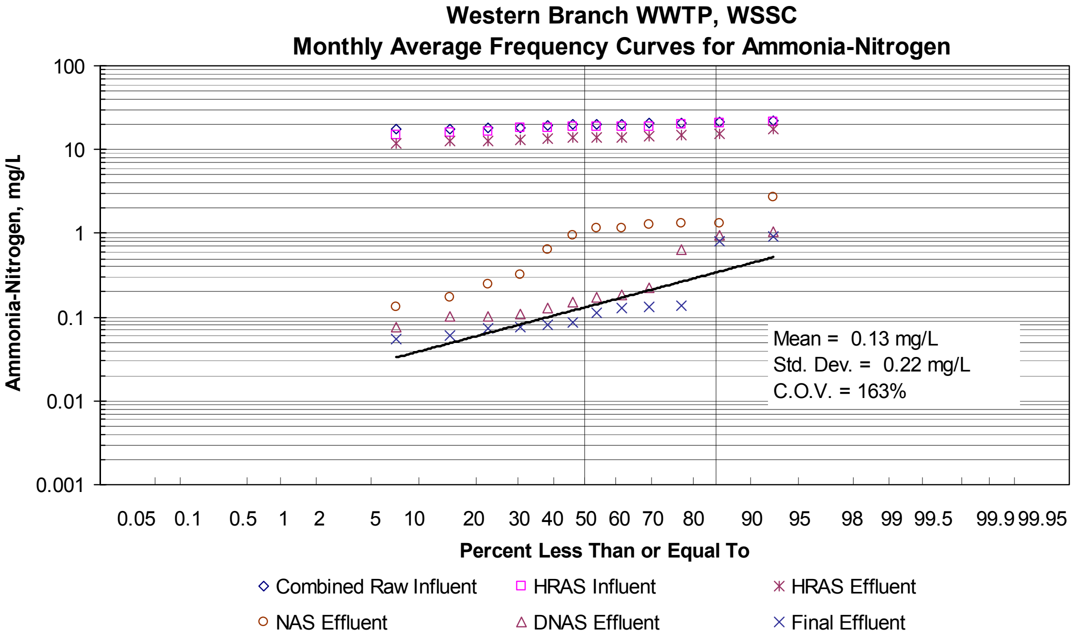

Figure 4. Monthly average frequency curves for ammonia nitrogen.

Figure 5 presents reliability data for the removal of total nitrogen (TN). The plant gives outstanding total nitrogen removal, with effluent TN of 1.63 mg/L and a COV of 36 percent.

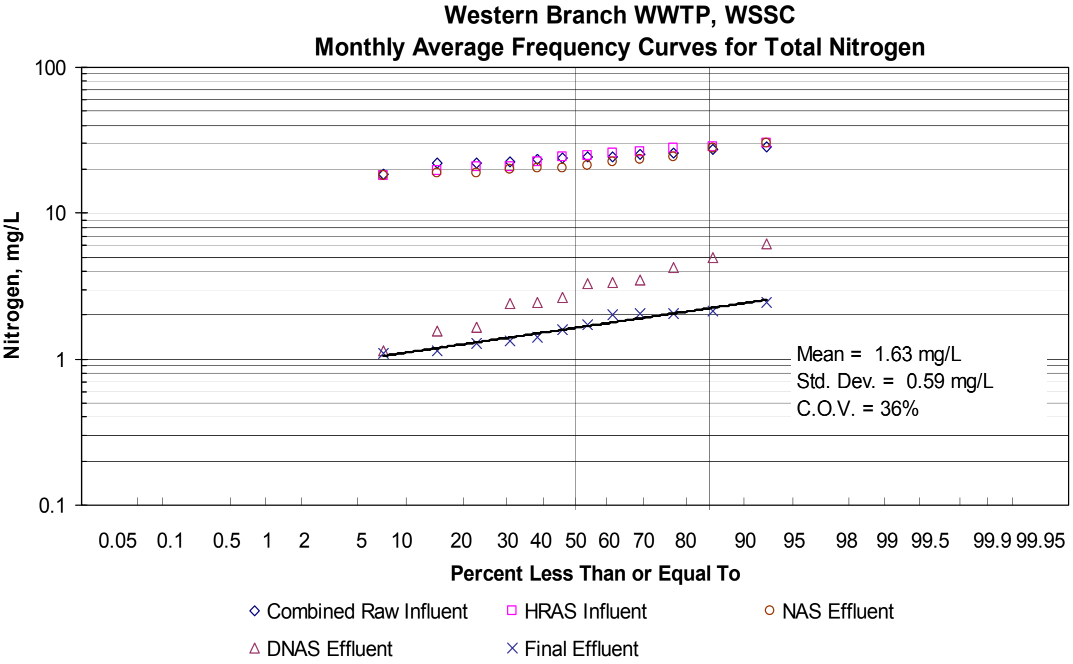

Figure 5. Monthly average frequency curves for TN.

Reliability Factors

This facility is unique in three ways: (1) three separate activated-sludge systems operated in series with dedicated clarifiers and RAS lines for biochemical oxygen demand (BOD) removal, nitrification, and denitrification with methanol feed; (2) chemical phosphorus removal; and (3) tertiary filtration. The facility also is unusual in that it has no primary settling. All sludge generated is biological and chemical sludge combined, which is incinerated after thickening by DAF and dewatering by centrifugation.

The results were excellent. The plant achieved a TN concentration of 1.63 mg/L with a COV of 36 percent and a TP concentration of 0.43 mg/L with a COV of 62 percent. Many factors accounted for this performance, and the key factors are presented below.

Wastewater characteristics: Because this facility uses a separate stage for denitrification, the use of an external carbon source (methanol) is a requirement. In addition, phosphorus removal is designed to be achieved with alum feed. The typical ratio for characterizing the adequacy of BOD is not applicable because the plant does not rely on internal carbon sources for biological removal of nitrogen or phosphorus.

The plant added a new process for nitrogen removal as the third step of the treatment train and called it DeNitrifying Activated Sludge, or DNAS, with separate clarifiers. The existing plant had a two-stage activated-sludge process before the current expansion: high-rate activated-sludge, or HRAS, for BOD removal and nitrifying activated sludge, or NAS. Both had separate aeration basins and dedicated clarifiers. The first two stages provided effluent with good BOD removal and full nitrification. The average concentrations in NAS effluent were 16.5 mg/L in nitrate-nitrogen with a COV of 12 percent and 1 mg/L in ammonia nitrogen. Note that nitrate-nitrogen is high because denitrification was not designed for. The third step, DNAS, proved effective in nitrogen removal. The control strategy included daily testing of key parameters, as well as adjustment of the dosage on an as needed basis. No online sensors are used in the DNAS basin.

A comparison of design vs. actual parameters follows:

Parameters	***Design***	***Actual***
HRAS HRT (hours)	2.68	3.0–3.8
HRAS Sludge age (days)		0.9–7.3
NAS HRT (hours)	5.51	6.7–7.9
NAS sludge age (days)		21–84
DNAS HRT (hours)	2.68	3.0–3.8
DNAS sludge age (days)	5–10	5–58

Another key feature of the plant is chemical phosphorus removal. Alum is added to the stripper/reaeration channel of the DNAS process at an average concentration of 10 mg/L and has proven effective. The tertiary filter is another key in providing reliability in nitrogen and phosphorus removal.

Methanol is added to the DNAS tanks at an average rate of 1,165 gpd to provide sufficient carbon for denitrification to occur. The methanol dosage is approximately 2.5 lb per pound of nitrate entering the DNAS tanks. Nitrate is checked by chemical testing three times a day to allow methanol dosage adjustment. The sludge generated is settled out with the rest of the DNAS sludge, mixed with the HRAS and NAS sludge, and thickened in the DAF units.

The facility employs online monitoring of dissolved oxygen (DO) in the HRAS and NAS basins, with one DO probe per reactor cell. The probe signals are used to control air valves and thus control the air feed to the basins. The plant also has online suspended solids probes in the aeration basins, which are used for monitoring, as well as sludge blanket monitors in the DNAS clarifiers.

Another key feature of the plant is that there is no primary settling. All sludge comes from the three biological systems, and the sludge is thickened aerobically at DAFs before dewatering and incineration. The recycle loads of nitrogen and phosphorus, therefore, remain low because there is no anaerobic digestion.

Wet-weather operation: Normal operating procedures are followed. No off-line storage is available.

Costs

Capital Costs

The plant was constructed in three phases. Phase 1, carried out in the early 1970s, included a dual sludge system for achieving BOD removal and nitrification, as well as filters that accomplish both nitrogen and phosphorus removal. The Phase 1 construction was sized for 15 MGD. Phase 2, completed in the late 1970s, consisted of additional tanks and filters to bring the dual sludge system to 30 MGD. Chemical addition for phosphorus removal was installed temporarily in the late 1980s, but not as a capital expense. Phase 3, carried out in the early 1990s, added a third sludge system for denitrification, along with making the alum addition system for phosphorus removal permanent. Table 5 shows the costs of those improvements, along with capital cost updates based on the *Engineering News-Record* Capital Cost Index (ENR CCI). The ENR CCI, which is compiled by McGraw-Hill, provides a means of updating historical costs to account for inflation, thereby allowing comparison of costs on an equal basis. From a Web site provided by the U.S. Department of Agriculture (USDA 2007), the ENR index for 1973 was 1,895; for 1976, 2,401; for 1991, 4,835; and for May 2007, 7,942.

Table 5. Plant improvement costs

	Year	Original cost	2007 cost	%P	%N	% other	Phosphorus cost	Nitrogen cost
Phase 1	1973	$15,000,000	$62,865,435	0%	20%	80%	$0	$12,573,087
Phase 2	1976	$7,500,000	$24,808,413	0%	30%	70%	$0	$7,442,524
Phase 3	1991	$30,000,000	$49,278,180	5%	60%	35%	$2,463,909	$29,566,908
Total			$136,952,028				$2,463,909	$49,582,519

The table also shows the percentage of capital cost for each phase that was attributed to phosphorus or nitrogen removal; the rest of the capital cost was attributed to other treatment, particularly BOD and TSS removal. Because the plant does not do biological phosphorus removal, it was assumed that only 5 percent of the Phrase 1, 2, and 3 costs could be attributed to phosphorus removal, which is a portion of the costs for filtration, plus the alum addition system. Nitrification was installed during both Phase 1 and Phase 2, but Phase 1 included additional activities not included in Phase 2, such as incineration and disinfection systems. Thus, 15 percent of the Phase 1 cost was attributed to nitrogen removal, whereas 30 percent of the Phase 2 costs were attributed to nitrogen removal. Since a large part of Phase 3 was the denitrification unit, it was assumed that 60 percent of the Phase 3 costs were for nitrogen removal.

The above analysis resulted in a total of $6,850,000 in capital attributed to phosphorus removal and $41,500,000 attributed to nitrogen removal, in 2007 dollars. The annualized capital charge for phosphorus removal (20 years at 6 percent) was $598,000. The annualized capital charge for nitrogen removal was $3,620,000.

The total capital attributed to nutrient removal, in 2007 dollars, was $48.4 million. For the 30-MGD facility, this means the capital expenditure per gallon of treatment capacity was $1.73.

Operation and Maintenance Costs

The plant uses chemical phosphorus removal and biological nitrogen removal, with extensive use of alum for the former and methanol as a supplemental carbon source for the latter. This means that the cost for phosphorus removal is essentially all chemical and for the disposal of the resulting sludge, with a small amount of electricity; the cost for nitrogen removal is electrical (for the aeration basins), chemical for the methanol, and for the disposal of the extra sludge resulting from methanol addition. A summary of the electrical calculations is provided in the Attachment. When the average electrical rate of $0.10/kWh (including demand charges) was applied, the cost of electricity for nitrogen removal was $229,000.

The average amount of alum applied for phosphorus removal over the period was 14.4 gallons per MG of flow, or 502 tons; at a cost of $212.25/ton, the cost of alum was $106,000. This cost was entirely attributed to phosphorus removal.

Methanol is applied at the DNAS to promote nitrate removal. The total amount of methanol added over the study period was 425,000 gallons. At an average cost of $1.00/gallon, the chemical cost for nitrogen removal was $425,000.

The alum added (9.5 mg/L as alum, or 0.86 mg/L as aluminum) was assumed to all convert to aluminum hydroxide sludge; at the average flow of 19.2 MGD, this was 400 lb of aluminum sludge per day, or 73 dry tons/year. The plant's average cost of disposal, considering trucking and incineration, was $440/dry ton. This made the cost of sludge for phosphorus removal $32,400.

The 425,000 gal/yr (2.8 million lb/yr) of methanol has a chemical oxygen demand (COD) of 1.5 lb COD/lb of methanol, or 4.2 million lb COD/yr. The typical yield of volatile suspended solids (VSS) on methanol is 0.4 lb VSS/lb of COD, giving 1.7 million lb sludge/yr, or 839 tons sludge/yr from methanol addition. At a cost of $440/dry ton, this made the cost of sludge for nitrogen removal $372,000.

Unit Costs for Nitrogen and Phosphorus Removal

During the evaluation period, the plant removed 213,000 lb of phosphorus. With the results above, the unit O&M cost for phosphorus removal is $0.78, while the unit capital cost is $1.01/lb of phosphorus removed.

During the evaluation period, the plant removed 1.32 million lb of total nitrogen. With the results above, the unit O&M cost for nitrogen removal is $0.99, while the capital cost is $3.27/lb of TN removed.

Life-cycle Costs for Nitrogen and Phosphorus Removal

The life-cycle costs are the sum of the unit capital and unit O&M costs. Thus, the life-cycle cost for phosphorus removal is $1.78/lb of phosphorus removed, and the life-cycle cost for TN removal is $4.27/lb of TN removed.

Assessment of magnitude of costs: The capital cost of $1.73 per gpd capacity is about average for the case studies. The capital for phosphorus removal is low, whereas the capital for nitrogen removal is high because of the use of the separate third stage for nitrogen removal. The O&M costs for phosphorus removal are low, whereas those for nitrogen removal are high because of the large amounts of chemical use with associated sludge generation.

Discussion

Reliability factors: This facility has a unique feature—three activated-sludge systems for biological treatment for nitrogen removal and chemical addition for phosphorus removal, followed by tertiary filtration. The reliability was excellent: the average concentrations were 1.63 mg/L in TN with a COV of 36 percent and 0.43 mg/L in TP with a COV of 62 percent.

For nitrogen removal, the third process, DNAS, relies on the external carbon source (in this case methanol), and the dosage was reasonable at 2.5 lb per pound of nitrate-nitrogen applied. The high level of nitrate in the NAS was noted. Chemical phosphorus removal was consistent in meeting the current limits.

Many factors have contributed to this reliable performance. The first key factor is the three separate processes in series—BOD and ammonia removal in the first two activated-sludge systems, followed by a separate activated-sludge system to denitrify with an independent supply of carbon. The fluctuations in wastewater or operating parameters and thus performance in one stage possibly can be balanced by the succeeding processes to achieve overall reliability in the plant's performance. An increased reliability for nitrogen removal was achieved through the use of an external carbon source; thus, the performance was not dependent on favorable wastewater characteristics. In addition, operating all four trains

(30-MGD capacity) while having a 19.3-MGD average influent flow contributed to excellent performance.

Note, however, that this unique system required a significant amount of land for aeration and clarification tanks; separate sludge return systems; and associated control equipment to operate, maintain, and monitor.

The cost for capital was low at $1.73 per gpd capacity as an upgrade. The O&M costs for phosphorus and nitrogen removal were $0.78/lb and $0.99/lb, respectively. The life-cycle cost for nutrient removal was $1.78/lb for phosphorus and $.4.27/lb for nitrogen.

Summary

The Western Branch WWTP is an advanced facility with a unique, multiple-system activated-sludge system followed by tertiary filtration. The facility was expanded and upgraded to meet new requirements with the maximum use of existing technologies. The latest upgrade included a third activated-sludge system for nitrogen removal, or DNAS.

The nitrogen removal was efficient and reliable at the mean concentration of 1.63 mg/L in TN with a COV of 36 percent. The phosphorus removal was also efficient and reliable at the mean concentration of 0.43 mg/L with a COV of 62 percent.

Many factors have contributed to this reliable performance. The first key factor is the three separate processes operating in series—BOD and ammonia removal in the first two activated-sludge systems, followed by a separate activated-sludge system to denitrify with an independent supply of carbon. The fluctuations in wastewater and/or operating parameters and thus performance in one stage were balanced by the succeeding processes to ensure the overall reliability of the plant’s performance. Performance was also enhanced by operating all four treatment trains (30-MGD capacity) while the influent flow was only 19.3 MGD. Phosphorus removal was achieved by adding chemicals to the DNAS.

Capital costs for the upgrade were low at $1.73 per gpd capacity. The O&M costs for phosphorus and nitrogen removal were $0.78/lb and $0.99/lb, respectively, and the life-cycle cost for nutrient removal was $1.78/lb for phosphorus and $.4.27/lb for nitrogen.

Key contributing factors for reliability include the inclusion of a separate third stage for denitrification. The separate stage with substantial methanol feed is able to provide a high degree of denitrification. That extra volume also provides further dampening of wastewater fluctuations, resulting in a very consistent effluent quality.

A separate stage for denitrification not only increases capital costs for the equipment but also necessitates the use of significant amounts of methanol to effect the needed denitrification. Phosphorus removal costs are reasonable with the use of alum for precipitation.

Acknowledgments

The authors acknowledge with gratitude the significant assistance and guidance provided by Robert Buglass, principal scientist for WSSC, and Nick Shirodkar, Plant Engineering supervisor at the Western Branch facility. This report would not have been possible without their prompt response with well-deserved pride in their facility and operation. EPA acknowledges the WSSC for its participation in this case study.

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Attachment: Electrical Costs

			kW		kWh	kWh	%P	%N	P kWh	N kWh
	HP	Number	Power draw	hours/ day	draw/day	draw/year				
HRAS/NAS blowers	1,500	1	1,119	24	26,856	9,802,440	0%	50%	0	4901220
Raw pumps	250	2	373	24	8,952	3,267,480	5%	20%	163,374	653496
Denite mixers	20	16	238.72	24	5,729.28	2,091,187.2	0%	70%	0	1463831.04
ID fans	75	1	55.95	24	1,342.8	490,122	0%	10%	0	49012.2
Final RAS pumps	100	4	298.4	24	7,161.6	2,613,984	0%	0%	0	0
Stripping channel blowers	200	2	298.4	24	7,161.6	2,613,984	0%	20%	0	522796.8
Centrifuge	300	1	223.8	24	5,371.2	1,960,488	5%	5%	98,024.4	98024.4
Air lift pump blowers	60	6	268.56	24	6,445.44	2,352,585.6	0%	50%	0	1176292.8
Total draw						25,192,270.8			261,398.4	8864673.24

Appendix B: Reliability, Variability, and Coefficient of Variation

When operating a treatment facility, the objective is to regularly produce an effluent that meets the discharge standards specified in the permit. Such regularity can be difficult to obtain because the measured effluent concentration of all constituents will vary. Some variations will be due to process upsets caused by weather conditions, accidents, and equipment failure. Others will be due to natural variations in influent conditions, as well as natural variability in laboratory measurements, sampling, and flow. In selecting a process, one possible criterion is finding one that has a higher probability of regularly producing a high-quality effluent and thereby keeps the facility well within permit compliance. The *reliability* reflects the overall performance of the facility in regularly meeting the treatment objectives, exclusive of extraordinary events like process upsets. Evaluating reliability or variability allows for screening of technologies by an assessment of how well a system might perform daily.

The variability of a data set can be represented by the coefficient of variation (COV). The COV is one standard deviation divided by the mean, expressed as a percentage.

Figure B-1 illustrates the meaning and determination of COV. By definition, a normally distributed population of data, such as measurements of total phosphorus in an effluent, results in a straight line when plotted on probability paper, as shown in Figure B-1. The mean of the data set falls at the 50 percent position, while one standard deviation from the mean can be found at plus or minus 34 percent, or at the 84 percent and 16 percent positions (McBean and Rovers 1998). This means that if the data are normally distributed, 68 percent of the results will have values within one standard deviation above or below the mean value. For the given period of evaluation, the slope of the line represents the reliability, or COV (i.e., the steeper the slope, the less reliable the performance; conversely, the flatter the slope, the higher the reliability). For example, Figure B-1, which shows effluent phosphorus for the Noman M. Cole Pollution Control Plant in Fairfax County, Virginia, indicates that the COV is 21 percent for the monthly averages for total phosphorus.

Note that the calculated reliability is a function of the data-averaging period. For the same year, COVs can be determined for the monthly average concentrations as well as the weekly average concentrations. In the example above, the COV of the Fairfax County facility is higher for the weekly averages, while the mean value is practically the same—29 percent on the weekly average, as compared to 21 percent on the monthly average. The same can be true with the COVs on the basis of a daily maximum at 45 percent.

For the purposes of this document, COVs are primarily based on monthly averages for consistent interpretation and easy comparison. When necessary because of the permit requirements, however, COVs with reference to the weekly averages are added. The decision to select a given averaging period is important and should be based on the permit conditions.

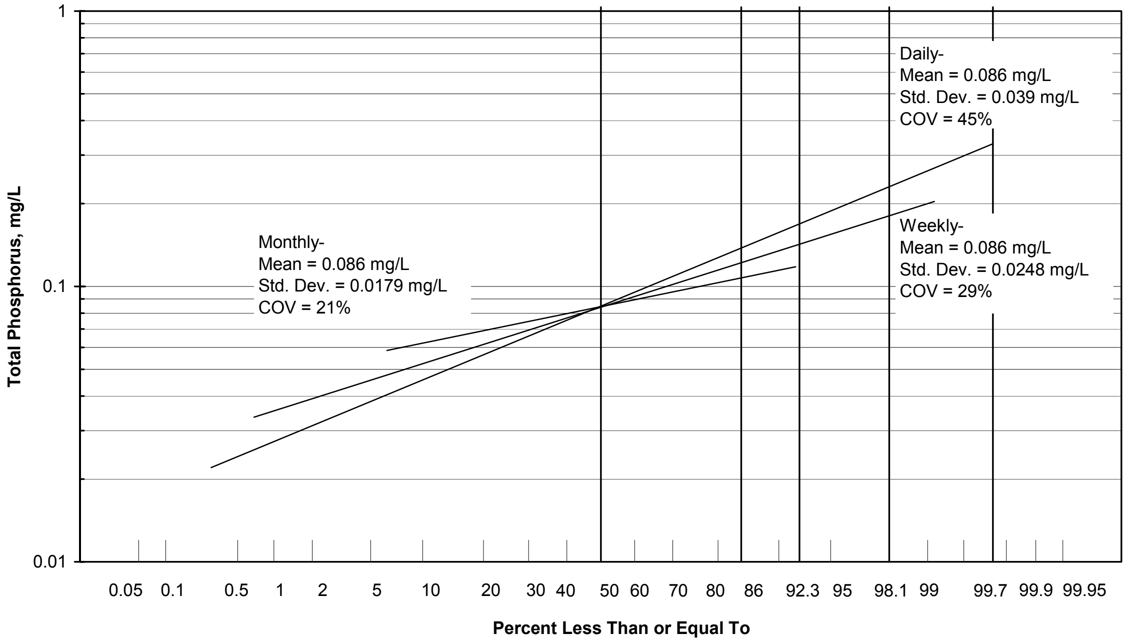

Figure B-1. Noman M. Cole Pollution Control Plant, Fairfax County, Virginia—daily frequency curves for total phosphorus.

The overall reliability of a facility increases with the increase in the number of processes installed in series, as shown in Figure B-2. For example, the reliability of a tertiary treatment facility would be higher than that of a secondary treatment facility. The designer of a facility can select multiple processes in series to increase the reliability of the entire treatment system. For example, the following data from the Noman Cole facility show total phosphorus concentration and COVs at each step of the treatment system:

- Secondary effluent: 0.74 mg/L at COV of 50 percent
- Tertiary clarifier effluent: 0.36 mg/L at COV of 33 percent
- Tertiary filter effluent: 0.09 mg/L at COV of 29 percent

The decision to add a particular level of reliability depends on the proposed permit limit and the degree of safety to be incorporated.

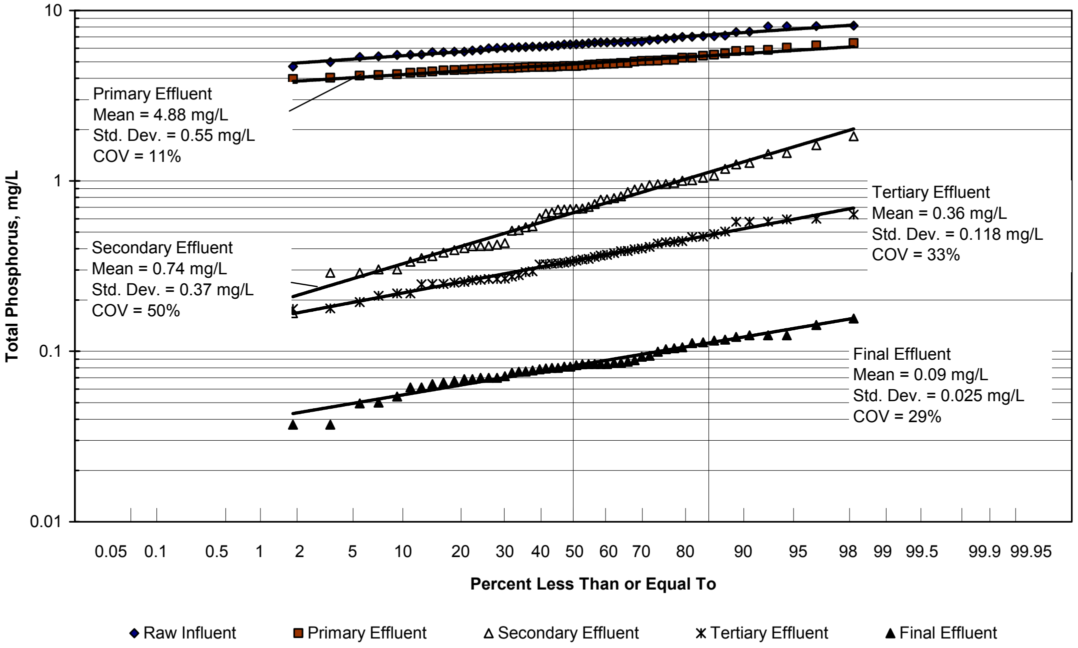

Figure B-2. Noman M. Cole Pollution Control Plant, Fairfax County, Virginia—weekly average frequency curves for total phosphorus.

Reference

McBean, E.A. and F.A. Rovers 1998. Statistical Procedures for Analysis of Environmental Monitoring Data and Risk Assessment. Prentice Hall PTR.