Alternatives to Chlorine — Is Ultraviolet Radiation the Answer?

Ann Farrell and John Burris

Widespread use of chlorination/dechlorination technology has recently come under increased scrutiny because of public health and safety concerns and potentially adverse air and water quality impacts. Consequently, there is a need to identify disinfection alternatives which are effective and economically feasible replacements for gaseous chlorine and sulfur dioxide. One disinfection alternative that shows great promise is ultraviolet radiation (UV), which eliminates many safety hazards and poses no threat to air or water quality.

Why Discontinue Chlorine Use?

The federal standards that are driving wastewater treatment plants away from chlorine to alternative disinfectants, such as UV, include modifications to facilities or new facilities associated with upgrading existing disinfection systems to comply with *Article 80 of the Uniform Fire Code (UFC) of 1988.* In addition to UFC, Federal OSHA standards for occupational safety and health apply.

Disinfection facilities must also operate in compliance with federal, state, and local air quality regulations. Potential air quality concerns resulting from disinfection facilities include emissions of air toxics (e.g., chlorinated organic compounds) and potential hazards from the accidental releases of toxic gases. The major regulations include the 1990 Clean Air Act—Title III Hazardous Air Pollutant (HAP) emissions standards and Risk Management Prevention Plan (RMPP) hazardous release requirements.

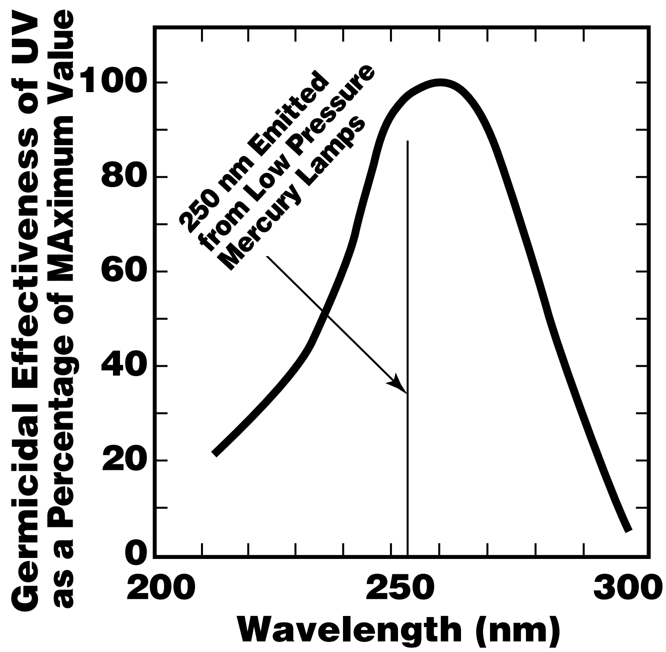
New water quality standards enacted EPA include stringent controls on chlorinated organics in effluents whose presence is increased by the use of chlorine. These standards may drive treatment plants to reduce their chlorine use. For all of these reasons, non-toxic alternatives to chlorine, such as ultraviolet radiation, should be given serious consideration.

Why Consider Ultraviolet Radiation?

Because UV uses only a light source, there is no potential for release of hazardous gases or for chemical reactions to form toxic byproducts. Disinfection by UV is a physical process that relies on the transference of electromagnetic energy from a source to an organism's cellular material. When UV energy is absorbed by the genetic material (DNA) of a microorganism, structural changes that prevent propagation of the organism are photochemically induced. The optimum wavelength for germicidal effectiveness is between 240 and 260 nanometers. UV disinfection facilities typically use low pressure mercury lamps, encased in long thin quartz tubes, as the UV source. The lamps produce nearly monochromatic light within this optimal wavelength range, and the tubes are contacted with the wastewater in basins similar to chlorine contact basins. The effectiveness of UV disinfection is a direct function of the intensity of the dose and the time of exposure (typically 5 to 20 seconds).

Considerations For UV Applications

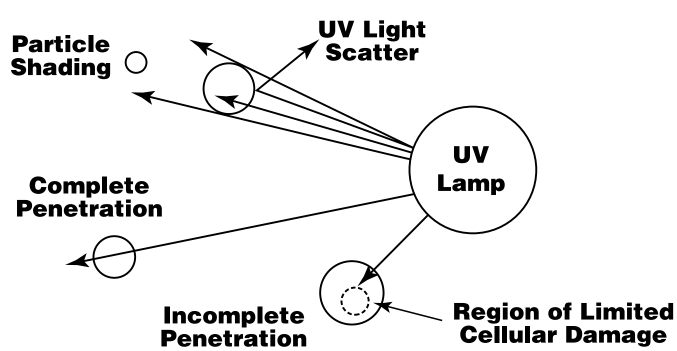
UV applications have several considerations. We would like to focus on two. First, the UV sizing is dependent on wastewater



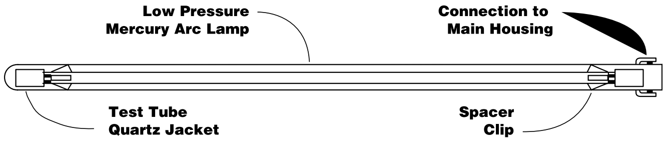
Typical UV Lamp Wavelength Spectrum

quality. Secondly, UV facilities can often be easily retrofitted into an existing facility. Because UV is a light source, its effectiveness is directly dependent on its ability to penetrate wastewater. Thus, two important determinants of UV effectiveness are wastewater turbidity and UV transmittance. Based on wastewater quality, both these parameters can vary widely depending on the types of dischargers to a given facility. In particular, UV transmittance is critical to the cost effectiveness of a UV application. Our research has shown that UV transmittance should be 50 percent or greater in order for UV to be cost effective. Interestingly, plants with industrial discharges of dyes or yeasts or plants using ferric chloride for odor control can have UV transmittance much lower than 50 percent, making these plants poor choices for UV applications.

A second key wastewater quality parameter is the fouling characteristics of the wastewater. As the UV lamps in their quartz sheaths become fouled, they lose their ability to emit UV light. Thus, a fouling factor must be assumed in the facility



Effects of Solids on UV Effectiveness



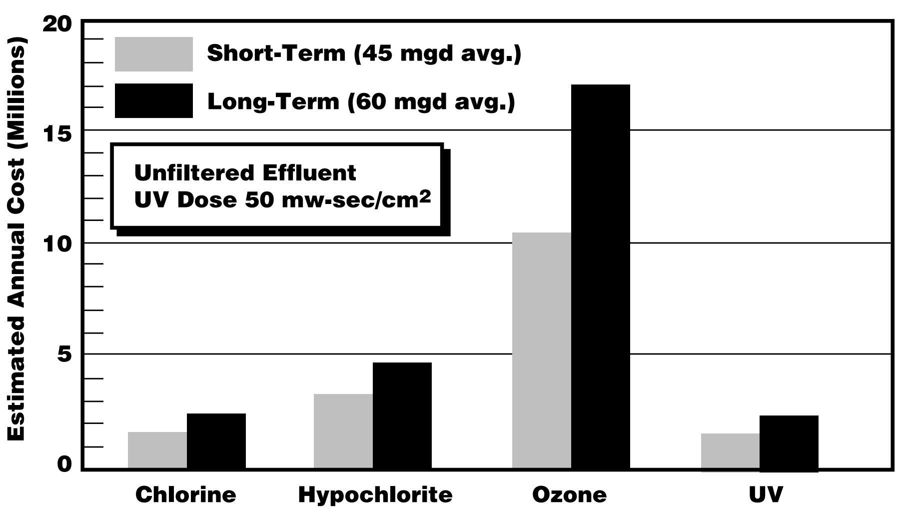
Typical Ultraviolet Lamp Assembly

design that accounts for the loss in effectiveness. Fouling is typically caused by alkaline precipitates and thus is a function of the pH and alkalinity of the wastewater. The assumed fouling factor can be manipulated to change the dose and number of lamps specified by increasing or decreasing the frequency of cleaning by acid washing. For facilities designed for frequent cleaning and reduced number of lamps, careful attention needs to be paid to the design of the cleaning facilities to provide for worker safety and ease of operation.

UV facilities require a much lower contact time than chlorine and can often be retrofitted into existing chlorine contact tanks. When this is attempted, careful attention must be paid to system hydraulics. The hydraulic grade line must be maintained within a two- to three-inch band for optimum UV performance. This necessitates the use of either a weighted flap gate on the discharge, where head is available, or a long weir where head loss is a concern.

What Other States Are Doing

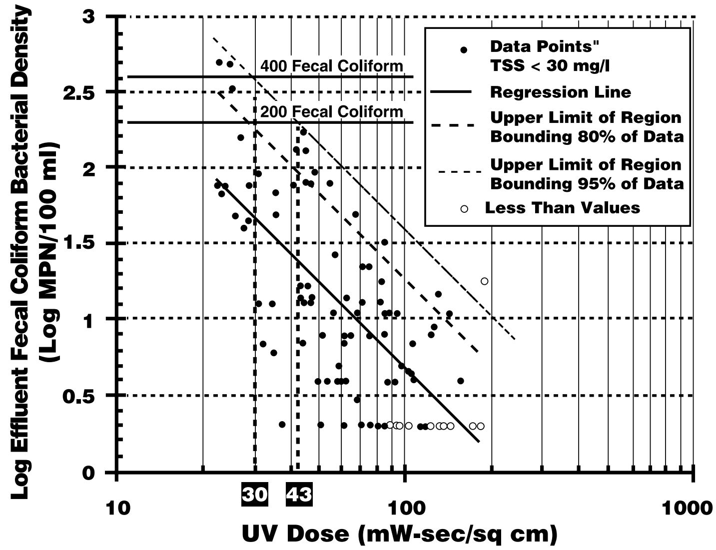
Montgomery Watson has investigated ultraviolet light and conducted intensive pilot testing for several California applications, including disinfection of secondary effluent discharged to a receiving water and disinfection of tertiary effluent for reuse. Pilot studies have demonstrated the equivalency of ultraviolet light to chlorine for reducing both coliforms and viruses. California's regulators are known for their conservatism, yet they have embraced the use of ultraviolet light based on the results of the pilot studies for both discharge and reuse. The pilot studies have generated data on required doses that have been used to size and cost facilities.



Estimated Total Annualized Costs of Disinfection Alternatives

The pilot studies and facilities planning analyses have shown ultraviolet disinfection to be cost competitive with chlorine in many situations.

Ultraviolet light has been an accepted disinfectant in many states for as long as ten years. UV equipment manufacturers estimate that there are currently over 25 UV installations in the nation, and the numbers are growing dramatically as more and more facilities look for alternatives to chlorine.



Horizontal-Lamp Pilot Unit
Effluent Fecal Coliform Bacterial Density Vs. UV Dose
For TSS<30 mg/l

How This Applies To Florida

Florida has a similar wastewater discharge situation to California in that a large percentage of its wastewater is reused. The rigorous approval process for ultraviolet radiation followed in California may benefit Florida dischargers in paving the way for the approval and use of ultraviolet radiation.

Basic disinfection in Florida is applied to secondary effluent and is defined as 200 fecal coliforms per 100 ml of sample. The geometric mean of 10 samples collected in a 30-day period shall not exceed 200 fecal coliform per ml. No more than 10 percent of the samples shall exceed 400 fecal coliform per 100 ml. No more than 10 percent of the samples shall exceed 400 fecal coliform per 100 ml. Using pilot plant data developed on a California facility with a minimal UV transmittance of about 50 percent would result in a clean lamp dose of 30 to 43 mw-sec/sqcm. Using a fouling factor of 80 percent, the design dose should be in the range of 38 to 54 mw-sec/sqcm. From our cost comparisons of numerous facilities, this is a cost effective UV operating range.

High level disinfection includes additional TSS control and fecal coliforms below detection. The comparable standard in California is filtration and total coliforms less than 2.2 MPN/100 ml. A rigorous review of UV technology by California's Department of Health Services and the National Water Research Institute using a panel of recognized experts resulted in the publication of "UV Disinfection Guidelines for Wastewater Reclamation in California." These guidelines specify an operational average dose for high level disinfection of 140 mw-sec/sqcm and various reliability requirements. If this dose is cost prohibitive, the discharger may elect to perform pilot testing and petition the Department of Health Services for a lower dose. Many facilities in California are installing UV disinfection in reuse facilities.

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Sequencing Batch Reactors for Nitrogen and Phosphorus Removal

Kenneth A. Mikkelson and Sara A. Ward

Prior to 1990, the city of Bonifay operated a trickling filter secondary treatment system designed to process 0.60 MGD of wastewater. Due to excessive infiltration and inflow and periodic equipment malfunctions, the flow-through treatment system was unable to meet DER and EPA permit requirements. The existing and projected flows from expanded service areas exceeded the discharge permit criteria. As a result, DER issued a new waste load allocation requiring an advanced level of treatment virtually making the trickling filter system obsolete. In July 1990, after the customary evaluation of treatment alternatives, system design, bidding and construction, Bonifay discharged the first effluent from its new 1.40 MGD design capacity sequencing batch reactor.

The SBR process was chosen for several reasons: land required for installation is minimal; capital costs are significantly less than a flow-through activated sludge system; microprocessor control means less operator attention than a comparably-sized flow-through system; SBR flexibility enables it to be operated for today's discharge requirements as well as future, more stringent requirements, usually without the addition of equipment or additional reactors.

Sequencing Batch Reactor —General Operation/Description

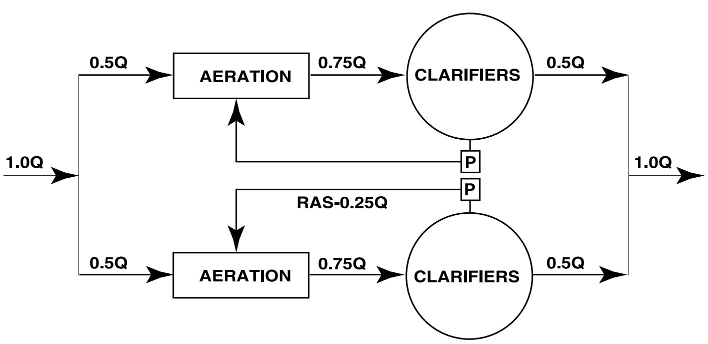
The sequencing batch reactor is an activated sludge process in which all treatment steps take place in a single reactor. In a flow-through activated sludge system, waste flows continuously through a series of basins composed of a minimum of aeration and secondary clarification (Figure 1). Discharge is continuous. In the SBR process, influent feeds continuously, but in batches, to each SBR reactor (Figure 2). In a two-reactor system, for example, 100 percent of the influent flows to each reactor half the time. While one reactor is receiving influent flow, the other reactor continues through the treatment phases. Discharge is intermittent.

The SBR system operates on the simple concept of introducing a quantity of waste to a reactor, providing a adequate time for the treatment of the waste, and subsequently discharging a volume of effluent plus waste sludge that is essentially equal to the original volume of waste introduced to the reactor.

This "fill and draw" principal of operation involves the basic steps of fill, react, settle, and decant. SBR systems are designed, within the overall fill and draw mode, to include seven individual phases of operation.

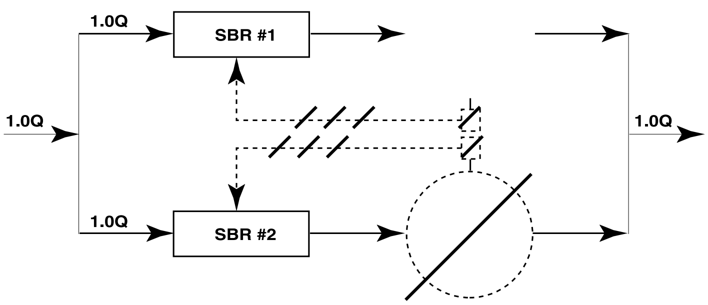
The phases are based upon process considerations related to effluent quality requirements of the specific wastewater treatment application. Phase duration is based upon specific waste characteristics and effluent objectives. The system at Bonifay employs an AquaSBR system. The AquaSBR reactor is equipped with mechanical mixing and aeration systems that operate on an independent basis. This capacity to mix only, without introducing oxygen to the reactor, or to mix and supply oxygen simultaneously, provides flexibility. By introducing or withholding oxygen at appropriate intervals, the operator can manipulate the reactor environment to create an anoxic or aerobic condition.

At the beginning of the process cycle, the SBR reactor is at its minimal liquid level and influent wastewater enters the reactor. Depending on the effluent required, the SBR will either be mixed in the absence of oxygen (Figure 3 "Mixed Fill") or mixed and aerated (Figure 3 "React Fill") or may be operated with alternate periods of mixing and mixing/aeration to achieve



FLOW: 0.5Q TO EACH HALF OF THE TREATMENT SYSTEM, 100% of TIME

Figure 1. Flow-Through Design



FLOW: 1.0Q TO EACH HALF OF THE TREATMENT SYSTEM, 50% of TIME.
SECONDARY CLARIFIERS, RETURN SLUDGE PUMPING AND PIPING NOT REQUIRED IN SBR SYSTEM

Figure 2. SBR Design

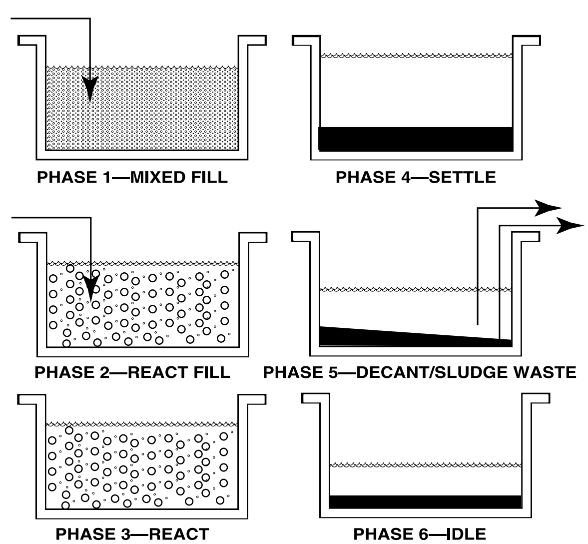


Figure 3. SBR Operating Phases

nitrification and denitrification. Mixing is provided by means of a floating mechanical downdraft mixer while aeration can by provided by any of several means, including membrane or coarse bubble, diffusers, or mechanical surface aerators.

In most SBRs, when the microprocessor (using preset timer values) or level sensors signal the reactor has reached the end of the React Fill Phase, influent flow stops and the basin enters the React Phase of treatment (Figure 3 “React”). Influent flow is diverted to a second SBR basin, where the treatment phases are initiated.

Following the React Phase, all equipment shuts off and the reactor contents are allowed to settle (Figure 3 “Settle”). Settling in an SBR occurs under perfectly quiescent conditions, since no flow is entering or leaving the reactor and no equipment is in operation.

When the sludge has settled for a predetermined period, the floating decanter weir will open and draw clear supernatant from 8-10 inches below the surface of the water (Figure 3 “Decant/Waste Sludge”).

When the Decant and Waste Sludge phases are complete, the reactor enters the Idle Phase (Figure 3 “Idle”) until the Fill phases are complete in reactor two. Influent is then diverted back to reactor one.

Phases in a Typical SBR System

Mixed Fill Phase: Just prior to the start of the Mixed Fill Phase, the reactor contents exist in a stratified condition, with the reactor environment having been “conditioned” by events that occurred during the previous cycle. The residual supernatant layer above the settled sludge mass is of reasonably good quality with respect to the concentration of specific waste parameters. The supernatant layer will contain BOD, COD, phosphorus, NH3-N and NOx-N, in concentrations at or near design effluent values.

As this phase begins, wastewater enters the reactor and the independent mixer begins operation, resuspending the settled biomass and bringing the microbial population in contact with the influent wastewater. No oxygen is introduced into the reactor. As organic and ammonia-nitrogen enter the reactor, the ammonia-nitrogen concentration increases. The increase of soluble organic material in the reactor, with the absence of dissolved oxygen, stimulates the release of phosphorus from certain organisms. The concentration of soluble phosphorus in the reactor will typically increase throughout the Mixed Fill Phase.

The reactor status during the Mixed Fill Phase is characterized by a completely mixed condition, with a uniform blend of biomass and supernatant from the prior cycle, plus the raw wastewater entering the reactor. It may be classified as anoxic with dissolved oxygen levels at zero and oxidized nitrogen at or near zero.

A treatment cycle structure incorporating this anoxic phase, regardless of the necessity for nutrient reduction, can be effective in controlling filamentous populations in the reactor. The use of anoxic conditioning of the sludge mass may be highly effective with respect to improved settling characteristics.

React Fill Phase: During the React Fill Phase, influent flow and mixing continue while aeration is added, converting the reactor from an anoxic to an aerobic environment. The effluent quality requirements, specifically with respect to the reduction of oxidized nitrogen, dictate that the aeration system will operate in a specific cyclical pattern. With denitrification as a requirement, the oxygen supply system is designed to operate in a specific “on-off” pattern. The capability to manipulate the oxygen supply source, in combination with the mechanical mixer, effectively creates alternating aerobic and anoxic reactor environments - resulting in efficient denitrification.

Ammonia-nitrogen concentrations in the reactor will steadily decline as the React Fill Phase is completed. Nitrification typically reduces NH3-N concentrations in the reactor as the raw waste flow continues to enter with additional organic and ammonia nitrogen.

Organic concentrations in the reactor, as evidenced by the soluble BOD or COD concentrations, will typically decrease as biological oxidation occurs simultaneously with the addition of organic material to the reactor during the React Fill Phase. The decline in BOD/COD concentration will closely parallel the pattern observed for the NH3-N concentration. The soluble phosphorus concentration within the reactor will sharply decline as the phosphates released during the Mixed Fill Phase are re-absorbed along with additional phosphates that enter the reactor.

React Phase: During the React Phase, influent flow to the reactor stops. Aeration may continue, or be cycled on and off, as in the React Fill Phase. Mixing also continues.

The importance of this phase should be recognized with respect to the opportunity to reduce concentration levels of all wastewater constituents without the influence of additional wastewater entering the reactor. In effect, the React Phase provides a time to “polish off” contaminants to the required concentration levels.

The removal of organic material as detected by a profile of the soluble BOD or COD concentration in the reactor, indicates a general decline from the initiation of aeration at the start of the React Fill Phase through the completion of that phase. The rate of decline in the React Phase by comparison, with the absence of any additional influent wastewater entering the reactor, is dramatically increased.

In summary, the React Phase features a completely mixed reactor which may incorporate intermittent or continuous aeration. As no flow enters the system, the phase provides an opportunity to “polish off” wastewater contaminants and reduce the concentration of organic material, ammonia and oxidized nitrogen.

Settle Phase: By this time, the preceding phases have accomplished all of the process objectives related to nitrification, denitrification, biological phosphorus removal reduction or organic compounds and conditioning of the biomass. During the Settle Phase no flow enters the reactor (aeration and mixing have ceased) and the reactor has become, in essence, a “static clarifier.” This perfectly quiescent environment is ideal for solids-liquid separation.

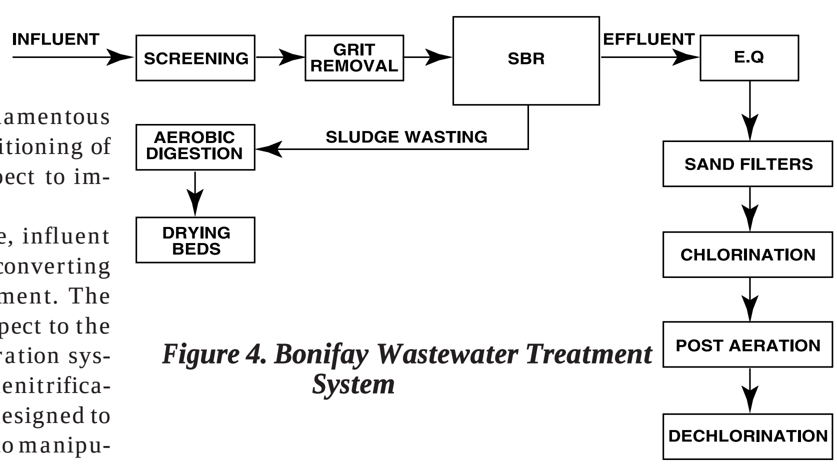


Figure 4. Bonifay Wastewater Treatment System

Decant Phase: Following the Settle Phase, it is necessary to remove approximately the same volume of liquid that entered the reactor during the Fill phases. This is achieved in the AquaSBR system during the Decant Phase, during which treated effluent from the upper part of the reactor is discharged by means of a floating AquaDecanter. The AquaDecanter is installed in such a way that it is able to rise and fall with changing water levels during the operating phases. A weir, located approximately 12 inches below the surface of the reactor, opens up - allowing the treated effluent to be removed, while avoiding the inclusion of any floating debris which could contaminate the final effluent.

Idle Phase: The Idle Phase may or may not be employed in an AquaSBR system and is a variable time period. Its duration is contingent upon specific hydraulic aspects of the treatment system.

Waste Sludge Phase: SBR systems, like other activated sludge process variations, are dependent upon a mixed culture of bacteria and other microbial life forms to accomplish treatment objectives. As a result of the biological degradation of organic matter and the accumulation of inert material present in most wastewaters, it is necessary to discharge certain quantities of solids to maintain an appropriate concentration of mixed liquor suspended solids in the reactor. This phase of operation within the treatment cycle is designed as a time increment that may occur after or simultaneously with the Decant/Idle Phase.

Bonifay WWTP Case History

A schematic of the Bonifay wastewater treatment system is shown in Figure 4. Following screening and grit removal, the wastewater is pumped to the SBR reactors. The effluent then flows to equalization basins and to sand filters. Filter effluent is chlorinated, aerated, and dechlorinated before discharge into a tributary of Branch Creek. Sludge is pumped from the SBR basins to aerobic digesters, then dewatered on drying beds. Dried sludge disposed of in a local landfill.

The SBR system at Bonifay utilizes two 87 by 87-foot square reactors with a sidewater depth that varies from 11.0' to 16.5'. The reactors are equipped with coarse-bubble diffused aeration, 100 HP positive displacement blowers, one 30 HP AquaDDM mixer and dual gravity AquaDecanters. The system was designed to treat 1.40 MGD of wastewater with a BOD of 175 mg/l, TSS of 175 mg/l, TKN of 40 mg/l, and phosphorus of 10 mg/l. Design parameters and discharge requirements are shown in Table 1.

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Table 1. Design Parameters

	Influent	Effluent Requirement
Flow	1.40 MGD avg. 2.50 MGD max. daily	
BOD	175 mg/l	8 mg/l
TSS	175 mg/l	8 mg/l
TKN	40 mg/	I5 mg/l (total nitrogen)
Total P	10 mg/l	3 mg/l

Table 2. Aqua SBR Performance Data—1991

	Influent						Tertiary Effluent					
DATE	FLOW	MAX DAY	CBOD5	TSS	TKN	T-P	CBOD5	TSS	TKN	NOx-N	Total-N	T-P
1991	MGD	MGD	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Jan	0.830	1.646	104	110	11	6.7	3.0	2.0	1.9	2.7	4.6	4.0
Feb	0.921	1.541	74	71	19	5.0	4.0	3.0	1.4	3.6	5.0	3.1
Mar	0.991	2.210	109	115	20	4.8	2.0	2.0	1.7	2.5	4.2	3.6
Apr	0.787	1.548	128	146	-	4.0	3.0	2.0	1.2	1.2	2.4	2.8
May	0.849	1.231	122	124	-	5.0	3.0	2.0	1.5	1.8	3.3	3.3
Jun	0.797	1.550	104	116	25	3.8	3.0	3.0	1.0	0.9	1.9	2.5
Jul	0.746	1.008	148	158	20	5.0	3.0	3.0	1.9	1.1	3.0	2.9
Aug	0.604	0.814	134	147	21	7.0	4.0	4.0	1.2	1.1	2.3	3.5
Sep	0.550	0.750	147	170	25	8.2	3.0	3.0	1.5	0.5	2.0	2.8
Oct**	0.447	0.605	137	131	45	4.7	3.0	3.0	2.4	0.8	3.2	2.4
Nov	0.438	0.933	127	166	-	4.8	3.0	3.0	2.4	1.5	3.9	1.0
Dec	0.443	0.709	141	172	-	4.8	5.0	3.0	3.7	2.0	5.7	1.9
Avg	0.700	1.212	123	136	23.3	5.3	3.3	2.8	1.8	1.6	3.5	2.8
Design	1.400	2.500	175	175	40.0	10.0	8.0	8.0	*	*	5.0	3.0

* No specific requirement
** Treatment cycle adjusted to enhance biological phosphorus removal

Table 3. Aqua SBR Performance Data—1992

	Influent						Tertiary Effluent					
DATE	FLOW	MAX DAY	CBOD5	TSS	TKN	T-P	CBOD5	TSS	TKN	NOx-N	Total-N	T-P
1992	MGD	MGD	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Jan	0.645	1.239	129	139	-	3.8	3.0	3.0	2.3	2.8	5.1	1.4
Feb	1.017	1.966	73	83	-	2.5	3.0	3.0	2.0	2.5	4.5	1.4
Mar	0.936	1.546	91	80	-	-	2.0	2.0	2.1	1.9	4.0	0.9
Apr	0.660	0.788	107	89	-	-	2.0	3.0	1.6	1.2	2.8	1.2
May	0.550	1.287	124	228	-	-	2.0	6.0	2.0	0.6	2.6	1.9
Jun	0.994	1.883	110	90	-	-	2.0	3.0	1.2	0.4	1.6	0.9
Jul	0.896	1.472	151	113	-	-	3.0	4.0	1.0	0.7	1.7	0.5
Aug	0.968	1.914	102	84	-	-	3.0	2.0	0.8	1.2	2.0	1.0
Sep	0.911	1.878	147	112	-	-	2.0	2.0	0.9	1.4	2.3	1.0
Oct	0.657	1.080	204	131	-	-	4.0	2.0	0.5	1.3	1.8	0.9
Nov	0.955	2.208	121	88	-	-	2.0	2.5	0.4	1.7	2.1	1.7
Dec	0.805	1.298	110	100	-	-	2.0	2.0	0.5	1.4	1.9	0.9
Avg	0.833	1.547	122	136	-	3.2	2.5	2.9	1.3	1.4	2.7	1.1
Design	1.400	2.500	175	175	40.0	10.0	8.0	8.0	*	*	5.0	3.0

* No specific requirement

Table 4. Aqua SBR Performance Data—1993

	Influent						Tertiary Effluent					
DATE	FLOW	MAX DAY	CBOD5	TSS	TKN	T-P	CBOD5	TSS	TKN	NOx-N	Total-N	T-P
1993	MGD	MGD	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Jan	1.117	1.941	125	58	-	-	2.7	2.2	0.3	1.0	1.3	1.2
Feb	0.905	1.203	206	109	-	-	2.5	2.0	0.4	1.8	2.2	1.1
Mar	1.112	2.024	125	108	-	-	3.0	2.0	0.5	1.1	1.6	1.5
Apr	0.786	1.190	118	158	-	-	1.8	1.7	0.5	1.1	1.6	1.4
May	0.586	0.766	209	185	-	-	2.0	1.7	0.4	1.9	2.3	1.0
Jun	0.487	0.561	206	252	-	-	2.0	1.8	0.3	1.7	2.0	1.3
Jul	0.554	0.788	215	240	-	-	2.3	2.2	0.4	1.1	1.5	1.8
Aug	0.583	0.808	166	206	-	-	2.0	1.7	0.4	0.3	0.7	1.3
Sep	0.615	0.876	159	174	-	-	1.8	2.0	0.4	0.5	0.9	1.7
Oct	0.494	1.574	136	182	-	-	1.9	1.9	0.7	2.4	3.1	1.3
Nov	0.574	0.772	169	150	-	-	3.0	1.2	0.5	0.8	1.3	0.7
Dec	0.646	1.194	142	118	-	-	1.9	1.7	0.5	1.1	1.6	0.9
Avg	0.705	1.141	165	162	-	-	2.2	1.8	1.3	1.2	1.7	1.3
Design	1.400	2.500	175	175	40.0	10.0	8.0	8.0	*	*	5.0	3.0

* No specific requirement

Effects of Water Treatment Plant Residuals on Advanced Wastewater Treatment Operations: Dunedin's Experience

Janice K. Magdziasz and Steven J. Duranceau

Wastewater flows by gravity to the Dunedin advanced wastewater treatment plant. From the wet well, raw sewage can be diverted to a 1.2 MG flow equalization tank and returned to the wet well when the raw sewage flow decreases. The plant is a 6.0 MGD A20 facility having effluent limits of 5 mg/l CBOD, 5 mg/l TSS, 3 mg/l Total N. and 1 mg/l Total P under both DEP and EPA discharge permits. The plant achieves high level disinfection through chlorination followed by dechlorination using sulfur dioxide. Aluminum sulfate (alum) is provided as a backup for biological phosphorus removal. Denitrification filters are augmented by methanol feed. The dissolved oxygen concentration of the effluent must be maintained above 6.0 mg/l. Oxygen is supplied by blowers using a manifold shared by the A20 aeration tanks.

The wastewater plant upgrade and expansion construction project was completed August 2, 1991. Since some grant money was received for construction, the city was required to submit a report to DEP summarizing the first year of operations. Close monitoring of process control was critical for a successful first year. The wastewater operators were still learning how to optimize the operation of the anaerobic/anoxic/aeration system, the denitrification filters, and chemical feed systems in 1992 when the new water treatment plant began operation. The city's reclaimed water system, 1992 pumping about 0.5 MGD, was placed into operation on August 4.

Dunedin's membrane softening water treatment plant is a 9.5-MGD rated facility utilizing pressurized green sand filtration with continuous injection of potassium permanganate, sulfuric acid addition, antiscalant addition, and micron filtration as pretreatment processes. The spirally wound membrane softening units are mounted in four independent skids that are designed to achieve 85 percent recovery. Post treatment consists of degasification, chlorination, fluoridation, and pH adjustment with sodium hydroxide.

The filter backwash water containing the iron precipitant is discharged to a 132,000 gallon holding tank for sedimentation. The supernatant is decanted from the tank and recycled to the head of the pressure filters. The backwash sludge is pumped into the city's sanitary sewer system.

A portion of the green sand filter effluent is directed to the membrane softening units through 5-micron filters, and the remainder of the effluent flows through 20 micron filters to blend with permeate from the MS units. Before MS treatment, the pH of the feed water is reduced and antiscalant is added to prevent scaling of the membrane softening units. The 15 percent of the MS feed water that does not pass through the membrane is disposed as concentrate through a 16-inch pipeline to several possible discharge points in the sanitary sewer system, which include a direct blend with the plant discharge and several other points within the wastewater treatment plant process.

The water plant startup was completed in phases. The green sand filters came on line June 8, 1992, only ten months after the last startup phase of the wastewater treatment plant. The MS process began regular operation on December 8, 1992. The water plant operators also went through a learning period to achieve effective process performance. The startup of both the green sand filter and MS processes effected the operations of the wastewater treatment plant processes.

Green Sand Filter Startup

Potassium permanganate is added to the influent of the pressurized green sand filters for the oxidation, precipitation and removal of iron, manganese, and hydrogen sulfide. The filters are designed to be backwashed every 24 hours or when the pressure loss across the filters reaches 10 psig.

On June 12, 1992, four days after the green sand filter startup, the chlorine demand at the wastewater treatment plant increased and was difficult to satisfy at times. The DO concentration in the reaeration tanks dropped from a norm of 6.0-8.0 mg/l to a range of 3.6-4.6 mg/l. The clarity in the dechlorination tank remained, but the effluent had a brown tint. The walls of the chlorine contact, reaeration, and dechlorination tanks were stained a reddish brown color by June 13.

By June 14, it was necessary to increase from one to two blowers to satisfy the oxygen demand in the aeration tanks and reaeration tanks. Operators began observing rust colored raw sewage in the grit removal tank for up to a one-hour duration.

The chlorine analyzer used for pre-dechlorination samples went out of service on June 16 and was followed by the post dechlorination chlorine analyzer on June 18. A black precipitant that was visibly suspended in the dechlorination tank was accumulating in the analyzers, fouling the elements and rendering them useless until they could be cleaned and returned to service.

Samples of the green sand filter backwash sludge, raw sewage and final effluent were taken June 16 to be analyzed for total iron, total manganese, sulfate and sulfite. Table 1 illustrates the results of these analyses.

On June 16, the alum feed was decreased from 40 to 20 ml/min to see what effect the filter backwash sludge was having on phosphorus removal. There was no significant change in phosphorus removal efficiency, so the alum feed rate was left in this pump minimum output mode.

Table 1. Green Sand Filter Impact On Wastewater Treatment Plant

	Water Treatment Plant Backwash Sludge 6/16/92 Grab	Raw Wastewater 6/16/92 Grab	Final Effluent 6/17/92 Composite
Iron, Total	19.1 mg/l	6.81	0.05
Manganese, Total	21.9 mg/l	8.02	0.12
Sulfate	31.9 mg/l	28.00	44.00
Sulfite	BDL	BDL	BDL

Table 2. Wastewater Treatment Plant Effluent

DATE	08/91	09/91	10/91	11/91	12/91	01/92	02/92	03/92	04/92	O5/92	06/92
FLOW MGD	4.146	3.868	3.707	3.593	3.547	3.580	3.643	3.714	3.520	3.428	3.477
BOD mg/l	1.8	2.2	2.0	1.8	1.5	1.8	2.1	1.9	2.4	3.0	2.0
SS, mg/l	0.7	1.1	1.0	1.7	1.4	2.0	1.9	1.7	2.3	2.0	1.0
TOTAL N. mg/l	2.33	2.27	1.94	1.96	2.56	3.72	2.23	1.96	1.94	1.54	1.74
TOTAL P. mg/l	0.43	0.29	0.37	0.44	0.35	0.68	0.28	0.30	0.33	0.29	0.32

DATE	07/92	08/92	09/92	10/92	11/92	12/92	01/93	02/93	03/93	04/93
FLOW MGD	3.734	3.825	4.065	4.151	3.689	4.034	4.334	4.433	4.676	4.252
BOD mg/l	3.0	1.3	1.0	1.0	2.0	2.0	2.0	2.1	2.2	2.7
SS, mg/l	1.0	1.0	1.0	1.0	1.0	1.0	1.0	1.3	1.3	1.6
TOTAL N mg/l	1.07	1.04	1.14	1.51	1.14	1.01	0.77	0.85	0.95	1.54
TOT AL P mg/l	0.15	0.24	0.14	0.21	0.28	0.15	0.17	0.11	0.17	0.30

By June 25 the city seal at the bottom of the dechlorination tank was no longer visible. Effluent suspended solids were increasing but were still within permit limits. An intense sampling and analysis program was established to determine the oxidation state of the iron, manganese, and sulfur through the plant processes. The results provided expected information without surprises. The iron and manganese were concentrated ten-fold from the raw sewage to the aeration tanks. The sulfates in the final effluent were greater than in the raw sewage due to our sulfur dioxide addition.

The effluent suspended solids returned to normal after a week. The remaining parameters had returned to near normal values with slightly discolored effluent through process acclimatization nearly two months after the green sand filter startup. The DO had returned to normal levels, and the chlorine demand had declined.

Membrane Process Startup

Dunedin's membrane softening water treatment plant was formally dedicated on September 15, 1992. However, the contract required 30-day membrane operation test did not begin until December 8. Between September and December the membranes were operated for brief periods.

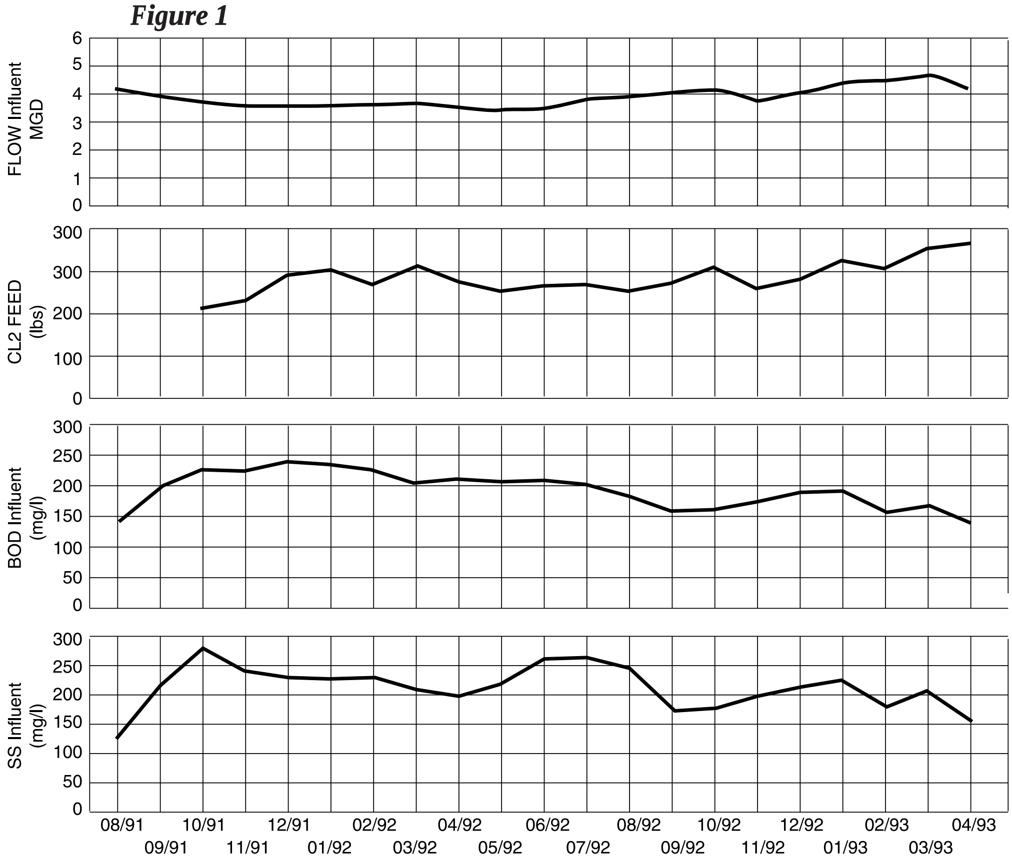
Sulfuric acid is added to the membrane feed water to lower the pH from approximately 7.3 to 5.8 to prevent the precipitation of sparingly soluble salt in the membrane. Antiscalant, a patented polyacrylic acid, is added to the feed water at a dose of 3 mg/l.

Membrane pretreatment also includes filtration designed to remove 90 percent of the 5-micron or greater sized particles. Each of the four membrane skids is designed to produce a maximum of 1.8 MGD of product water. Anticipated concentrate flow to the wastewater treatment plant was 0.5-0.75 MGD at a blend ratio of 13 20 percent.

Due to illness of a laboratory employee, phosphorus samples from late December through early January were held for analysis. When the samples were analyzed, the results indicated that the effluent total phosphorus had jumped from <0.5 mg/l to 5 mg/l or greater. The process field test kits had not indicated any problems. Concentrate and antiscalant were investigated as sources of additional phosphorus or interferences with the vanadate test method in use. It was determined that both the concentrate and antiscalant interfered with the vanadate method, consequently a conversion to the ascorbic acid method for phosphorus analysis was initiated. The December monthly average for effluent phosphorus concentration was 0.15 mg/l.

The water plant concentrate caused the wastewater plant raw sewage flow to increase from an average 3.711 MGD to 4.346 MGD or 17 percent. The increased flow served to dilute the CBOD and TSS in the raw sewage. The chlorine demand increased by 17 percent and stayed there. These parameters are displayed in Figure 1.

Figure 1



The DO demand increase in the aeration tanks caused three blowers to operate rather than two. The operators tried to maximize the output of two blowers before adding a third blower. Even during early morning, a low demand period typically, three blowers were needed. By mid January, two blowers were run again by compromising the DO setpoint for several hours each day.

While investigating the antiscalant as a method interference, it was found that its primary function is as a dispersant. The antiscalant also dispersed the activated sludge floc in the clarifiers resulting in turbid supernatant and solids washout. This increased the loading to the denitrification filters which caused additional backwashes and bumps to ensue. The activated sludge system solids inventory was increased and the system adapted within two months, the bumps were decreased and the backwashes remained the same.

The effluent CBOD increased from an average of 57 lbs/day (2.0 mg/l) before the startup to 85 lbs/day (2.2 mg/l) after the startup. The effluent pH decreased from an average of 7.3 to an average of 6.9 from January, 1992 through April, 1993.

The operators from the water and wastewater plants have worked together to communicate pending activities and to devise better ways to handle routine operations. The wastewater plant maintains a higher solids inventory, diverts stronger wastes to the flow equalization tank and closely monitors permitted parameters. The water plant provides notification of changes in operation and slowly feeds cleaning solutions to the sewer system.

Conclusions

As can be seen in Table 2, the wastewater treatment plant has always been well below required discharge parameters. Since Dunedin's customers are almost all residential, the acclimation of the process to this nontypical waste stream was a challenge. The following is a summary of our observations:

(1) Wastewater Treatment Plants can operate with discharges from MS processes long as waste streams are not released in slug loads and waste stream parameters such as pH and chlorides are reasonable.

(2) Although chlorine demand and oxygen demand decreased during filter startup, they returned to more normal levels within two months.

(3) The iron sludge seemed to cause a brown tint to the wastewater effluent and an increase in TSS. The TSS returned to normal in the effluent within one week and the color gradually returned to a more clear color after several months of operation.

(4) The ascorbic acid method for phosphorus analysis is appropriate for testing waste streams containing MS concentrate.

(5) Startup of the membranes caused an increase in flows of about 17 percent which diluted the CBOD and TSS. The city is looking into the possibility of permitting the direct blend of the concentrate with the wastewater plant effluent since the future treatment of brackish water may increase chlorides and impact the city's reclaimed water system. The piping and valves are already in place to accomplish this bypass.

Acknowledgments

Appreciation is extended to Gerald Knippel and Evelyn Bolin of the Dunedin membrane softening water treatment plant for their assistance, and to Robert Brotherton, P.E., Dunedin public works and utilities director, for his support in the preparation of this paper.

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Sequencing Batch Reactors from Page 36

Annual influent and effluent data for the period from January, 1991 to May 1994 are shown in Tables 2-5. A summary of influent and effluent data for the 41 months of operation from January, 1991 to May, 1994 is shown in Table 6.

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Table 5. Aqua SBR Performance Data—January-May 1994

	Influent						Tertiary Effluent					
DATE	FLOW	MAX DAY	CBOD5	TSS	TKN	T-P	CBOD5	TSS	TKN	NOx-N	Total-N	T-P
1994	MGD	MGD	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Jan	0.752	1.640	127	105	-	-	1.8	1.6	0.7	1.6	2.3	1.0
Feb	0.912	1.287	94	118	-	-	2.2	1.6	1.0	0.9	1.9	0.7
Mar	0.937	1.661	101	129	-	-	1.8	1.8	0.6	0.7	1.3	0.7
Apr	0.856	1.410	82	99	-	-	1.5	1.6	0.8	2.0	2.8	1.0
May	0.644	0.893	111	133	-	-	2.1	2.0	1.3	1.1	2.4	1.3
Avg	0.820	1.378	103	117	-	-	1.9	1.7	0.9	1.3	2.1	0.9
Design	1.400	2.500	175	175	40.0	10.0	8.0	8.0	*	*	5.0	3.0

* No specific requirement

Table 6. Aqua SBR Performance Data—January 1991-May 1994

	Influent						Tertiary Effluent					
DATE	FLOW	MAX DAY	CBOD5	TSS	TKN	T-P	CBOD5	TSS	TKN	NOx-N	Total-N	T-P
	MGD	MGD	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l	mg/l
Avg	0.755	1.310	133	134	23.3	5.000	2.6	2.4	1.1	1.4	2.6	1.2
Design	1.400	2.500	175	175	40.0	10.0	8.0	8.0	*	*	5.0	3.0

* No specific requirement