Anion Exchange Technology for Organic Control in Potable Waters

Troy L. Lyn and Steven R Lavinder

An anion exchange system will be used as an additive treatment process to an existing lime softening water treatment plant process to reduce color and meet the Stage I Safe Drinking Water Act Amendments Disinfectant/ Disinfection Byproduct (D/DBP) Rule. The anion exchange system has been piloted at a lime softening plant for an industrial facility in South Florida. A full-scale system has been designed and is being constructed. While membranes were considered, the lower capital and operation and maintenance cost of the anion exchange processes as an additional process to the existing lime softening process was more cost effective.

The water treatment plant currently uses chlorine oxidation to remove natural organic matter followed by air stripping to remove trihalomethanes (THMs) formed from the chlorine oxidation process. The natural organic matter is a source of color and a source of precursors of disinfection byproducts (DBPs) like THMs, which are becoming more stringently regulated by the D/DBP Rule. In addition, the new Stage I D/DBP Rule will expand the DBPs regulated to include haloacetic acids (HAAs) that are non-volatile and are not removed by the air stripping process.

Existing Water Treatment Process

In the existing treatment process, chlorine solution produced from chlorine gas is used to oxidize color and natural organics. The THMs formed during the chlorination process are removed by aeration. HAAs, which are also formed during the chlorination process, are not removed by volatilization. Following aeration, the water is lime softened. Chlorine solution is added in the settling basins of the softening unit to prevent algae growth in the basin. After lime softening, the water is sand filtered and then sent to the clearwell. Ammonia, added in the clearwell, forms chloramine to prevent the additional formation of THMs and other DBPs. The existing process uses a large quantity of chlorine to oxidize the color and natural organics. Approximately 400 lb/day of chlorine is used to treat 1 MGD of water. While the current process provides water with THMs below the current standard of 100 µg/L, the existing process will not be able to reduce THMs below the new THM standard of 80 µg/L. In addition, the existing process will not able to meet the new HAA standard of 60 µg/L.

Anion Exchange

To reduce color and meet the anticipated THM and HAA requirements, anion exchange was evaluated as an alternative by conducting a pilot study. The anion exchange process was considered after small demonstration studies at several other southern Florida utilities showed it was effective at reducing natural organic matter. Anion exchange resins are commonly used in industrial applications to remove organic concentrations in the production of high purity water and decolorization of food components such as syrup and molasses. Applications in potable water systems are limited because of the difficulty in disposing of the regenerate, variable organic removal efficiency depending upon the natural organic matter chemistry, and the potential for irreversible organic fouling if the resin is not properly selected or regenerated.

Typically, anion exchange installations for potable water systems are at small community systems. At this facility, the resin was evaluated for the reduction of color-causing organics and reduction in TOC and possible DBP precursors. The anion exchange resin used in the pilot study was a macroporous, strong base Type I styrenic resin (Thermax, A72-MP). The resin used was not virgin; it had been used in pilot studies at other south Florida water treatment facilities. Table 1 summarizes the physical characteristic of the styrenic resin.

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Table 1 Resin Characteristics

Functional Group	Quarternary Ammonium Type 1
Skeleton	Cross-Linked Polystyrene
Hydophobicity	Phobic
Pore Structure	Porous
Temperature Stability (°F)	195
Moisture (%)	58
Particle Size (mm)	0.3 to 1.2
pH Range	0 to 14
Mean Pore Radius (nm)	4,000 – 8,000

Pilot Study Conclusions

The anion exchange process was piloted using raw water and lime and filtered lime-softened water. The major conclusion of the study was that the anion exchange resin was very effective at removing color and exhibited the ability to reduce the HAA formation potential. Settled water color was reduced from approximately 25 to 30 color units to less than 1 color unit, and the raw water color was reduced from approximately 70 color units to less than 5 color units. The haloacetic acid formation potential was reduced in the raw water from an average value of 173 µg/L to less than 20 µg/L, and total organic carbon (TOC) was reduced from 11 mg/L to 1.1 mg/L. The pilot demonstrated that the anion exchange resin can successfully remove organics in the raw and filtered lime softened water. In summary, the anion exchange was able to do the following:

- effectively remove over 90 percent of color
- reduce total organic carbon by 50 to 60 percent
- reduce THMs and HAAs by 40 to 60 percent when free chlorine is used.

New Water Treatment Process

Based on the pilot study results, a new treatment scheme was developed. It involves adding ammonia and sodium hypochlorite to the raw water to form chloramine residual. Sodium hypochlorite was chosen as the new form of chlorine to eliminate the potential release of chlorine gas. The chloramined water is then aerated to remove excess carbon dioxide and add dissolved oxygen before lime softening. After lime softening and filtration, the water then passes through the anion exchange units before going to distribution or storage. Based on full scale testing with the chloraminated water, the chlorine demand was reduced to 100 lb/day. Regenerate from the anion exchange process will be sent to an industrial waste plant on site. However, the regenerant may have been diluted with other process water and sent to the domestic wastewater system.

Although the resin worked effectively to reduce color on the raw water, applying it before lime softening would reduce the life of the resin because of heavy organic loading. To maximize the life of the resin and to reduce the frequency of resin regenerations, it is desirable to locate the resin at a point that the organic loading will be lowest onto the resin. In this case, the desirable location is after the filtration process. The resin can be oxidized by a free chlorine residual so it must be applied in a location where no free chlorine will be in contact with it. To

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Comparison of Hollow-Fiber and Spiral-Wound Membrane Technologies on a Brackish Groundwater

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Since Sarasota placed into operation its existing hollow-fiber membrane process in 1982, advances in membrane technology have continued. As part of its *Water Supply and Treatment System Improvements Program*, the city desired to explore opportunities available in membrane technology.

The primary objective of the pilot plant phase of the investigation was to compare the performance and operation of advances in hollow-fiber membrane technology with spiral-wound membrane technology. The factors used to evaluate membrane performance included the effects on salt rejection, permeate production, recovery, pressure, flux, and membrane differential pressure. Cost and performance could then be compared between spiral-wound and hollow-fiber technologies. To support the evaluation, a multi-unit pilot plant was designed and constructed for comparing hollow-fiber and spiral wound technology. The membranes remained in operation for about six months.

Water Treatment Facility Description

Sarasota's water treatment facility is composed of two major water treatment plants: a reverse osmosis water treatment plant (ROWTP), and an ion exchange (IX) water treatment plant. The 12-MGD capacity WTF results from a combination of 4.5 MGD from the ROWTP and 7.5 MGD provided by the ion-exchange plant, of which 2.3 MGD is a blended bypass water.

The city's RO system is supplied by a network of eight wells. The RO system is permitted to withdraw 6.0 MGD on average and up to 6.5 MGD on a peak month basis. The wells pump into a common manifolded well piping network which feeds into the RO system.

Pretreatment for the RO system consists of the addition of sulfuric acid and scale inhibitor to the raw water stream, which is then sent through 1-micron rated cartridge filters. Following pretreatment, the raw water feed stream was historically boosted in pressure by three vertical aluminum-bronze turbine pumps (during the implementation of the study these pumps were replaced with stainless steel pumps). A fourth high-pressure pump is also available and serves as a backup. The operating pressure of the membrane process is approximately 390 psi.

The ROWTP, placed on-line in 1982, was originally configured as a high-pressure system (390 psi) utilizing DuPont B-9 polyaramid hollow fiber membrane assemblies. Design recovery efficiency is nominally 75 percent, which means that 6.0 MGD of raw water is required to produce the 4.5 MGD of product. The city's reverse osmosis system is composed of three separate sections or trains, each designed to produce 1.5 MGD of permeate from 2.0 MGD of raw water at a recovery of 75 percent and a total plant capacity of 4.5 MGD. All trains have the capability of operating independently from the others, and each train is composed of two stages. The remaining 1.5 MGD of concentrate (brine) must be discharged as a waste stream subject to DEP regulation.

Initially, each of the three process trains were comprised of 54 membrane bundles in the first stage and 21 in the second stage using 10-inch diameter DuPont B-9 permeator bundles. In August 1997 the city remembraned the first and second stages of Train C with a 36:18 array of DuPont Permasep B-9 Twin Permeators, which offered a more efficient bundle. At start-up, Train C permeators feed pressure was established at

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325 psi to produce approximately 1,000 gpm of permeate water. The initial recovery was set at 70 percent. The initial differential pressure was measured as 33 psi for stage 1 and 35 psi for stage 2.

In December 1997, Train B was also remembraned with a 36:18 array of DuPont Permasep B-9 Twin Permeators. Train A continued to rely on standard DuPont B-9 permeators that had been installed in 1995. In late 1997 DuPont announced its new product line of low pressure, 8-inch diameter cartridges that were available in double and triple internal staging configurations. As a result, the city initiated an investigation of DuPont's newest, ultra-low pressure double cartridge configuration that included a side-by-side comparison to the latest spiral-wound technology available.

Pilot Plant Description

Hollow-Fiber Pilot Plant. Multiple DuPont HF Cartridge Assemblies were used in the investigation. The HF cartridge assembly was designed for low pressure brackish water applications utilizing a high surface area/low flux membrane bundle with flexibility in system design at high recovery. The hollow-fiber pilot plant's hydraulic configuration includes a feed by-pass line in the brine bushing, allowing a portion of the feed water of one cartridge to bypass its bundle and pass directly to the next cartridge. This reduces flow through the first in a series of cartridges and avoids a high flow resistance, allowing the bundle to operate at lower pressures than the original bundle technology offered by DuPont. The HF cartridge can be inserted into traditional 8-inch diameter pressure vessels used frequently in SW applications.

Eight-inch diameter double cartridges were used to simulate a two stage membrane array. Two double cartridge assemblies were installed in parallel on Train C. Each double cartridge assembly was monitored independently of one another. The cartridge assemblies were supplied with feed water tapped directed from the high pressure feed header of train C.

Spiral-Wound Pilot Plant. The spiral wound pilot assembly was designed as a two-stage array, 2:1 configuration, with three spiral wound elements per pressure vessel. The spiral wound pilot plant was also designed to satisfy the requirements of the SDWA Information Collection Rule (ICR). The membrane elements used were 4 x 40 inch, TFC 4821 ULP, manufactured by Fluid Systems of Venice. Table 1 provides information relative to the two different membrane technologies.

Table 1. RO Membrane Technology Comparison

Membrane Type	HF Technology	SW Technology
Membrane Model	BW-L-8540, (double)	TFC® 4821 ULP
Nominal Membrane Area	4,475 ft^2 per cartridge	80 ft^2 per element
Nominal Permeate Flow	6.5 gpm	1.0 gpm
Minimum Concentrate Flow	3 gpm	4 5 gpm
Design Feed Pressure	100 –400 psi	50 - 175 psig
Maximum Feed SDI	5 units	5 units
Design Recovery	75% per cartridge	10 - 15% per element

Operation and Monitoring

Raw water from the downtown wellfield was used as the supply water for the HF cartridges and SW pilot plant. Sampling of the raw water was used to assist in the investigation and to provide a means for future membrane system analysis (modeling) and planning. The raw water sample was collected as a composite sample of the water entering the ROWTP from the wellfield. The wells that were operating at the time that the sample was collected were noted and recorded. A raw water sample was collected once a month over a three month period.

Table 2 lists some of the raw water quality parameters measured during the pilot-testing period. The raw water quality remained relatively consistent and was comparable with available historical data. The full-scale ROWTP relied on SHMP for scale control and sulfuric acid addition for pH adjustment prior to cartridge filtration. Although most pilot-scale studies contain integrated pretreatment, it was decided to demonstrate HF and SW technology on full-scale ROWTP pretreated water. This was primarily due to the fact that the full-scale system has previously experienced fouling; consequently, pilot testing under actual feed water conditions may provide insight into the full-scale performance decline.

Operation of the pilot plants was from September 1997 to February 1998. The pilot plants operated continuously with the exception of 112 hours of down time for train C that was dedicated to cleaning and other miscellaneous operation activities. The total run time for Cartridge Assembly No. 1 was 8660 hours. Cartridge assembly No. 2. operated for 8676 hours. The spiral wound pilot plant was operated a total of 2828 hours.

Table 2.Raw Water Analysis

Parameter	Median	Standard Deviation	Range
Turbidity (ntu)	0.04	0.03	0.02 - 0.08
Alkalinity (mg/L as $CaCO_3$)	130	1.0	129 - 131
pH	7.14	0.01	7.13 - 7.15
Total Organic Carbon (mg/L)	2.0	0.14	1.84 - 2.11
Total Iron (mg/L)	0.01	0.0	< 0.005 – 0.006
Total Hardness (mg/L as $CaCO_3$)	1202	72.1	1125 – 1269
Bromide (mg/L)	0.11	0.09	0.06 – 0.24
Total Suspended Solids (mg/L)	1.70	0.98	< 0.2 – 1.7
Total Dissolved Solids (mg/L)	2210	26.5	2200 – 2250
Sulfate (mg/L)	1000	41.6	996 – 1070
Chloride (mg/L)	352	2.52	349 – 354
Fluoride (mg/L)	1.20	0.06	1.1 – 1.2
Strontium (mg/L)	21	12.7	0.05 – 23
Silica (mg/L)	23	2.31	19 – 23
Magnesium (mg/L)	122	3.1	118 - 124

Water recovery, feed pressure, and membrane productivity were monitored during the pilot operation. Initially, water recovery was maintained at a constant level by varying the feed and concentrate flow rate. However, throughout the testing of the spiral-wound and hollow-fiber pilots, water recovery was varied between 50 and 75 percent for the HF technology, and between 50 and 70 percent for the spiral wound technology.

The pilot membrane systems (i.e. both HF cartridges and the spiral wound pilot) were allowed to operate for a minimum of 48 hours after start-up before water quality samples were collected. Field measurements of pH, temperature, flow, and pressure were taken frequently during that initial period as a means of determining steady-state conditions. After the cartridges and pilot plant achieved steady state operation, pH, temperature, and conductivity were measured and recorded every eight hours (once per shift) in the feed, permeate, and concentrate streams. Field measurements of flow and pressure (including interstage) were recorded every four hours.

Water Productivity

At the start-up of Cartridge Assembly No. 1, a feed pressure of 225 psi produced an average permeate flow of 15 gpm at a water recovery of 63%. Cartridge assembly No. 2 start-up conditions were very similar with a feed pressure of 225 psi and permeate flow of 16 gpm at a water recovery of approximately 65%. The spiral wound pilot plant was initially set at a feed pressure of 95 psi and a combined permeate flow of 14 gpm at

50 percent water recovery. Figures 1 through 3 presents feed pressure versus run-time as a function of water recovery increment for each pilot component.

The permeate flow for Cartridge Assembly No. 1 was normalized to a feed pressure of 235 psi, differential pressure of 7 psi, permeate pressure of 15 psi, and osmotic pressure of 40 psi. These values represent average data collected between September 15 and 26, 1997. The normalized permeate flow for Cartridge Assembly No. 1 ranged between 3 and 6 gpm for a recovery of 50%. However, an increase in feed pressure to approximately 330 psi produced a permeate flow ranging between 15 and 29 gpm. At a 55% recovery, the normalized permeate flow ranged from 23 to 36 gpm. When recovery was maintained at 63%, normalized permeate flow ranged between 9 and 17 gpm. An increase in recovery to 70% increased the normalized permeate flow to an average of 22 gpm.

Cartridge Assembly No. 2 was normalized during the same time period. The standard conditions consisted of a feed pressure of 236 psi, differential pressure of 6 psi, permeate pressure of 15 psi, and osmotic pressure of 42 psi. The normalized permeate ranges for Cartridge Assembly No. 2 for a recovery of 65% were between 13 and 16 gpm. An increased recovery up to 70% increased the normalized permeate to an average of 23 gpm. At 75% recovery, the normalized permeate flow was documented to have increased between 21 to 37 gpm.

In comparison, the ranges of normalized permeate flow for the spiral wound pilot plant were between 6 and 10 gpm for 50% recovery and 15 to 19 gpm for 58% recovery. However, increasing the recovery to 65% obtained a range of 16 to 20 gpm. Maintaining a 65% recovery while reducing the flux reduced the normalized permeate flow to 10 gpm. This reduced flux also produced an average normalized permeate flow of 10 gpm for a 70% recovery. The normalized membrane permeate flows for the hours of operation are shown in Figures 4 through 6 for each pilot plant component.

Figure 1. Feed Pressure of Cartridge Assembly No. 1 vs. Run Time

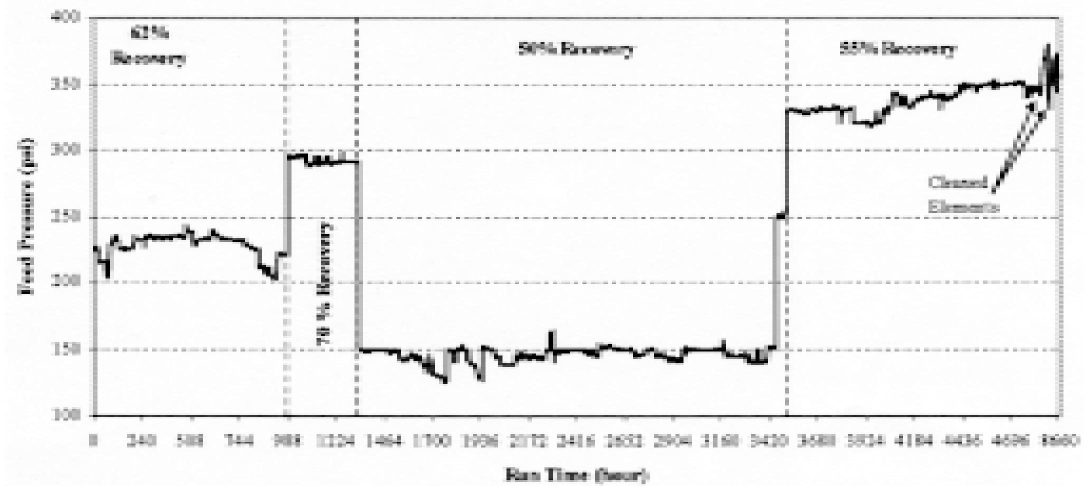


Figure 2. Feed Pressure of Cartridge Assembly No. 2 vs. Run Time

Figure 3. Feed Pressure of Spiral Wound Pilot Plant vs. Run Time

Differential Pressure

Cartridge Assembly No. 1 indicated fluctuations in differential pressure throughout pilot plant operations. However, after an increase in feed pressure, there appeared to be only a slight increase in differential pressure. Conversely, after approximately 2304 operational hours, Cartridge Assembly No. 2 indicated a gradual increase in pressure. At the end of pilot testing, the differential pressure for cartridge No. 2 was increased from 15 psi to 40 psi. The spiral wound pilot unit operated in a similar manner, with the differential pressure being increased from 14 to 49 psi at the end of pilot testing.

Figure 4. Normalized Permeate Flow – Cartridge Assembly No. 1

Figure 5. Normalized Permeate Flow - Cartridge Assembly No. 2

Figure 6. Normalized Permeate Flow - Spiral Wound Assembly

Salt Passage

Salt passage, as well as pounds of salt removed, were also significant parameters in the evaluation. Both parameters are a function of a calculated TDS concentration. Salt concentrations were calculated using the daily (sampled every eight hours) conductivity levels as measured from the feed, concentrate,and permeate water streams. The direct relationship between TDS and conductivity for all of the sample points was estimated as 0.86. In addition, salt passage was normalized to a salt correction factor of 93 for the cartridge assemblies and 23 for the SW.

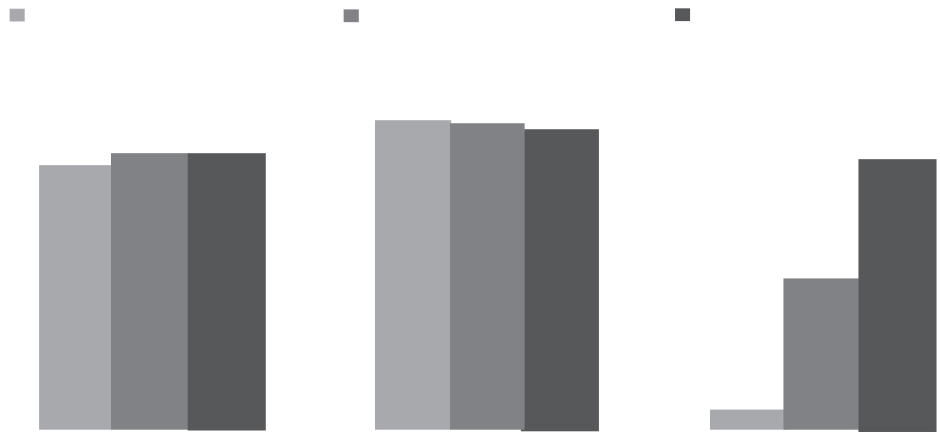
The salt passage for Cartridge Assembly No. 1 ranged from an average of 7% at a recovery of 50% to an average of 1.5% at a recovery of 55%. At 55% recovery, the feed pressure was increased in order to maintain an increased permeate flow. However, at 70% recovery, an average salt passage of 3% was obtained. Cartridge No. 2 operated similarly in that at 70% recovery an average of 3.8% salt passage was obtained, and at 75% recovery an average salt passage of 3% was achieved. The spiral wound indicated slightly improved performance for salt passage. At 50% recovery, the salt passage averaged 3.8%. However, at a decreased flux of 14 gsfd (70% recovery), the spiral wound salts passage averaged 4.8% in comparison to the 3.8% for the HF technology.

Water Quality Results

Figure 7 shows the results of the average TDS, sulfate and chloride removed for both HF cartridges and the SW pilot plant assembly. While the cartridge assemblies sufficiently removed TDS and sulfate, the performance of chloride removal was somewhat lower than the spiral wound. Cartridge assembly No. 1 showed a TDS removal of 96.7%, sulfate removal of 99.9%, and chloride removal of 81.3%. Cartridge assembly No. 2 showed a TDS removal of 97.7%, sulfate removal of 99.7%, and chloride removal of 89.7%. The spiral wound pilot plant resulted in a TDS removal of 97.7%, sulfate removal at 99.4%, and chloride rejection of 97.4%.

Mass balances were calculated for the cartridge assemblies and spiral wound pilot plant for various water quality parameters components. In addition, a two-sided two-sample *t*-test (alpha of 0.05) was performed to statically compare experimental data and theoretical predictions of incoming and outgoing mass to each membrane pilot unit. In all cases, the *p* values were within the 95 percent confidence levels, and there were insignificant percent differences between the input and outputs of the pilot plant assemblies. These results indicated very little accumulation of mass in the membrane modules for those conditions experienced during the pilot plant program.

Figure 7. Average TDS, Sulfate and Chloride Percent Removal



SW technologies sufficiently removed TDS and sulfate, the performance of chloride removal was significantly lower for the HF technology than the spiral wound membrane technology. The use of HF resulted in slightly more efficient sulfate removal than SW technology.

Based on schedule constraints, water quality needs, ease of installation, and other considerations, it was decided that cartridge HF would be used to retrofit the existing B-9 process Train A. Cartidge HF offered a very quick, reliable operating cost reduction method for the city over the traditional DuPont B-9 HF permeator configuration. This would enhance the city's near-term water supply goals, where longer-term goals that consider the conversion to spiral-wound construction could continue in a planned phasing approach, primarily due to more complex SW retrofit issues. Because the new cartridge HF technology would operate at half of the pressure than the original DuPont B-9 Twins, additional operation savings would be realized in the near-term by simply using HF cartridge technology that is conveniently offered in a tradition 8-inch diameter SW pressure vessel configuration.

Power Cost

The energy cost per kw-h was approximated at $0.075. The calculated power cost corresponds to the normalized permeate water produced. As an average, the energy cost for the hollow fiber technology at 70% recovery will be approximately $0.34/kgal. However, at 70% recovery, the energy cost for the spiral wound technology will be approximately $0.13/kgal. The spiral wound technology cost is lower due to the lower feed pressure requirements of each technology.

Conclusions

Both hollow-fiber and spiral-wound membrane technologies reliably produced high-quality treated waters relative to sulfate and total dissolved solids. Both technologies were capable of achieving 75 percent water recovery; however, pressure requirements for the hollow-fiber configuration was approximately double that of the spiral wound technologies. Power cost at 70 percent recovery was estimated at $0.34/kgal for hollow-fiber technology as compared to $0.13/kgal for spiral-wound technology. However, the cost to replace one skid with the new cartridge HF technology was $0.3 million versus approximately $1.1 million to retrofit one skid with SW technology. Also, the HF cartridge would operate at half the pressure regime than that of the original B-9 bundles.

Results indicated that the SW membrane appears to produce higher-quality water than the HF technology. While the HF and

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avoid this possibility, free chlorine oxidation, as practiced now through the settling basin and filters, will be discontinued. Instead, a chloramine residual will be applied before lime softening coagulation and carried through the entire treatment process. Chloramines do not adversely affect the resin and maintaining a chloramine residual of 5 mg/L is acceptable. Another benefit of eliminating the free chlorine oxidation process is that the formation of THMs and HAAs is minimized.

Anion Exchange System – Basis of Design

The anion exchange system is designed for the maximum rated capacity of the WTP. The system will consist of two treatment units, each with a capacity of 0.75 MGD for a total treatment capacity of 1.5 MGD. Two reactors are to be provided so that at least one of the units will always be in service. During periods that one unit is out of service for regeneration, the flow will be reduced or the water will be bypassed around the unit. The estimated time to backwash and regenerate one unit is approximately 2 hours.

The maximum flow through the resin system with one unit out of service will be approximately 1.0 MGD at a maximum loading rate of 3 gpm/ft^3. Plant operations will require that regeneration occur when water demands do not exceed that value. This corresponds to a hydraulic loading rate of 2 gpm/ft^3 of resin. If one unit is out of service for regeneration, the other unit can be loaded for a short term operation at 3 gpm/ft^3 for a total flow of 1.0 MGD.

The anion exchange vessels are sized with a diameter of 10 feet, which will require a resin depth of 3.33 feet (40 inches). Providing 120 percent resin expansion for backwash plus allowances for the top and bottom shells and distance above the floor, the total height of the unit will be approximately 14 feet.

In addition to the anion exchange vessels, a regeneration system is required. The regeneration process requires a saturated salt solution (approximately 26 percent or 10 lb/gal). The salt solution is obtained from a brine tank that can hold approximately 36 tons of salt. The salt will be loaded from a tank truck that will blow the salt into the tank. A water solution is then added through the tank to obtain the proper salt solution. The brine solution will be pumped to the resin vessels through a brine transfer pump. Regenerant brine that is used and disposed of is typically 3% of the treated water flow. Regeneration will typically occur in the following steps:

1. The batch product meter reaches a preset level and an alarm is sent to the control room indicating that a unit is ready for regeneration process.
2. The regeneration process begins with backwash of the resin with influent water. This cycle is typically an upflow process through the resin for approximately 7 to 10 minutes with a maximum length of 60 minutes.
3. Following upflow backwash, a brine solution is pumped into the unit in a downflow mode for approximately 20 to 25 minutes with a maximum of 40 minutes.
4. After the brine solution is applied, the strong salt solution must be rinsed from the resin by a slow rinse using influent water for 20-25 minutes. The time will be determined after the conductivity in the effluent drops to less than the maximum contaminant level for total dissolved solids. A fast rinse cycle can also be provided to further reduce the salt content before the units are returned to service.

Estimated Costs

The estimated capital cost for the anion exchange system alone is approximately $0.40/gpd. This assumes a 3-year resin life. The operating cost for the anion exchange system alone is estimated at approximately $0.08/1,000 gal. The operating cost for the anion exchange system with lime softening/filtration is approximately $0.48/1,000 gal. The operation and maintenance costs for the anion exchange system may vary considerably depending upon the frequency of resin replacement and regenerant disposal charges. The operation and maintenance costs will drop with longer resin life.

Summary and Conclusions

The anion exchange process is a feasible technology for removing organics from lime softened/filtered water. Because the treatment system had sufficient lime softening capacity to meet future water demands, replacing lime softening with membranes was not believed to be a cost-effective alternative to reduce color and meet the Stage I D/DBP Rule. The addition of the anion exchange process as an advanced treatment process to lime softening provided the following benefits:

- Reduced color below three color units on a consistent basis.
- Reduce chlorine demand of the treatment process.
- Reduce precursors to meet the Stage I DBP requirements. ■