Process Optimization at new Nanofiltration Plant Meets Water-Quality Goals, Saves Money for Boca Raton

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In recent years, the city of Boca Raton has experienced a gradual but steady increase in the levels of color in the raw and finished water at its Glades Road Water Treatment Plant. Also, high levels of dissolved organics in the raw water made compliance with new, more stringent regulations for disinfection byproducts (DBPs) more difficult with the plant's lime softening process. Faced with increasing customer dissatisfaction with high color levels in the finished water and stricter drinking-water regulations for DBPs, the city decided to install a 40-million-gallon-per-day membrane treatment process to reduce color and DBPs in its finished water.

Table 1 shows representative quality values of the raw-water supply from the city's three shallow (100 to 200 feet in depth) Biscayne Aquifer wellfields. This table summarizes the key permeate and blended finished-water quality goals that were the basis for the membrane process design.

Primary treatment objectives were to reduce finished-water color to <6 CU, which is essentially colorless to most consumers, and to reduce THM and HAA values in the finished water to meet the Stage 2 Disinfection/ Disinfection Byproducts Rule with a 20-percent safety margin. The permeate hardness range was set at 50 – 80 mg/l as $CaCO_3$, which would reduce hardness to a desirable range while retaining sufficient hardness to avoid corrosion concerns in the distribution system.

Early Pilot Testing

Early in the design phase of this project, the raw water at the Glades Road plant presented several challenges for developing a successful membrane treatment process. Initial membrane pilot testing revealed that cartridge filter and membrane fouling were significant problems. Fouling issues included sand and silt, oxidation of iron and hydrogen sulfide, biofouling, and dissolved organic foulants.

The extent of fouling encountered in the pilot testing made it evident that the raw-water quality must be improved through rehabilitation of the existing wellfield and/or enhanced pretreatment. Initial fouling of the cartridge filters was so rapid that they had to be replaced approximately every seven days. To address this problem, a multimedia filter was installed upstream of the cartridge filters to lengthen their run time.

After the multimedia filter was installed, the operating life of the cartridge filters improved significantly; however, membrane fouling continued to be a concern. The pilot plant operating data still showed a continuous decline in membrane flux due to dissolved organic foulants. These flux decline trends in all three stages of the pilot plant are shown in Figure 1.

A variety of pretreatment chemical combinations and membrane models were investigated in an effort to alleviate these fouling problems. Pilot testing activities aimed at addressing the problems included:

- Side-by-side comparisons of the effectiveness of various antiscalants and dispersants.
- Operation with just antiscalants or antifoulants without acid.
- Side-by-side comparisons of several membrane models with respect to fouling characteristics and hardness rejection.

Constituent/Parameter	Raw Water	Permeate	Finished Water
Total Hardness, mg/l $CaCO_3$	250	50 – 80	60 – 90
Alkalinity, mg/l $CaCO_3$	214	<175	<175
Total Dissolved Solids (TDS), mg/l	450	<300	<300
Color, CU	50	<2	<6
Total Organic Carbon (TOC), mg/l	12	<1	-
Humic Acid, mg/l	10	-	-
Total THMFP, mg/l	0.600	0.042	0.064
Total HAAFP, mg/l	0.400	0.030	0.048
pH	7.2	-	8.0 – 8.5

TABLE 1. TYPICAL RAW-WATER AND REQUIRED PERMEATE AND FINISHED-WATER QUALITY

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The results of these tests are summarized in Figure 1. During the operating period from 25,500 to 30,800 hours, four test runs using different combinations of acid and antiscalants were conducted. These tests included several brands of the most commonly used antiscalants and antifoulants in the industry. They were characterized by membrane fouling and rapid declines in membrane mass transfer coefficient. Fouling at these rates would require cleaning at less than three-month intervals, which was considered unacceptable.

Operation without Acid & Antiscalant

After several unsuccessful test runs with some of the most widely used pretreatment chemicals in the membrane industry, several membranes were removed from the pilot test unit for autopsy analysis. The results of this testing revealed that, while substantial fractions of the foulants were consistent with naturally occurring humic acids, other components consistent with the active ingredients of the antiscalant chemicals were also observed in significant concentrations.

The results of the membrane autopsies and observations from other membrane treatment plants in South Florida with high humic acid concentrations appeared to indicate that these antiscalants may actually be complexing with the naturally occurring humic acids and may be contributing to the fouling problem.

To test this theory, samples were sent to Dr. Harvey Winters of Fairleigh Dickenson University to conduct laboratory scale tests aimed at measuring the fouling potential of various membranes and antiscalants and dispersants in combination with Boca Raton's

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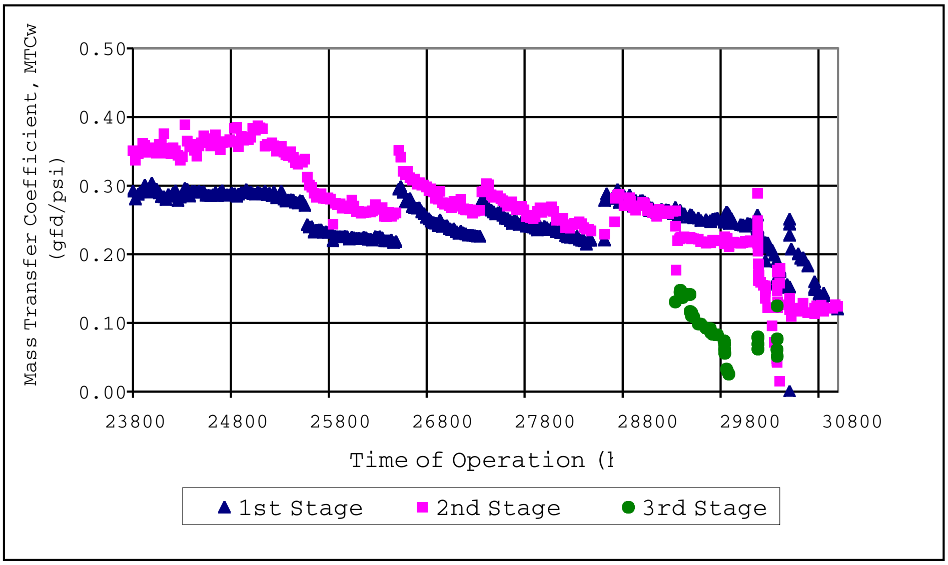

FIGURE 1. MASS TRANSFER COEFFICIENT VS. TIME OF OPERATION

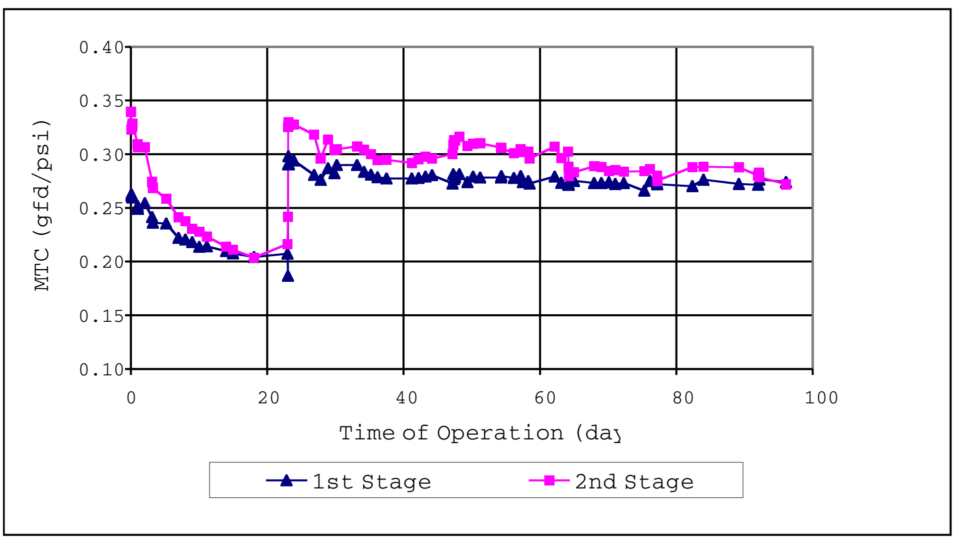

FIGURE 2. MASS TRANSFER COEFFICIENT VS. TIME OF OPERATION

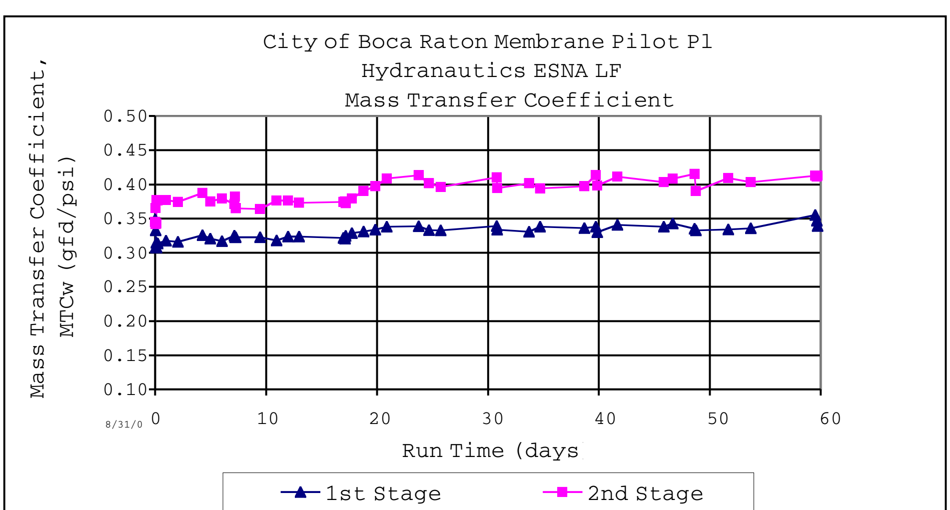

FIGURE 3. ESNA LF MASS TRANSFER COEFFICIENT VS. TIME

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feed water. Zeta potential measurements of the repulsion between negatively charged humic acids and the membrane surface charge were made and revealed significant differences between membranes. Most membranes showed a positive zeta potential at pH levels below 7, which would decrease the repulsion between humic acids and membranes and promote humic acid fouling.

Dr. Winters also performed electrophoretic measurements with polyacrylamide gels, which confirmed that some antiscalants and dispersants enhanced the adsorption of humic acids onto the membrane surfaces. Based on these test results, it was decided to operate the pilot unit without either acid or antiscalant.

A comparison of operation with and without antiscant is illustrated in Figure 2. The initial 20 days of operation is indicative of pilot unit performance using antiscalant with no acid addition. Operation during this period is characterized by fouling and a rapid decline in mass transfer coefficient in both stages.

Testing from day 22 through the remainder of the test run was conducted without acid or antiscalant pretreatment chemicals. Operation during this period was much more stable than in the previous testing with acid and/or antiscalant.

New Low-Fouling Membranes

Membranes used during this successful run without acid and antiscalant were a new prototype model from Hydranautics that were developed specifically to have a reduced fouling potential for waters with high humic acid concentrations. These membrane elements were predecessors to what would later become the ESNA LF line of low-fouling nanofiltration membranes. The ESNA LF membranes employ similar low-fouling technology as the LF series of membrane for low-pressure, reverse-osmosis (RO) applications.

The results of the first 60 days of testing with the new low-fouling membranes are shown in Figure 3. A minor adjustment was made to the pilot unit operation between day 17 and 20; however, it can be seen that the mass transfer coefficients during this period of operation were very stable. These results represented the best performance with respect to fouling that had been achieved with the pilot testing up to that point. The pilot test unit continued to operate stably with the ESNA LF membranes for approximately six to eight months without acid and antiscalant, and without cleaning.

Side-by-Side Testing

After satisfactory operating performance had been demonstrated without acid and anti-

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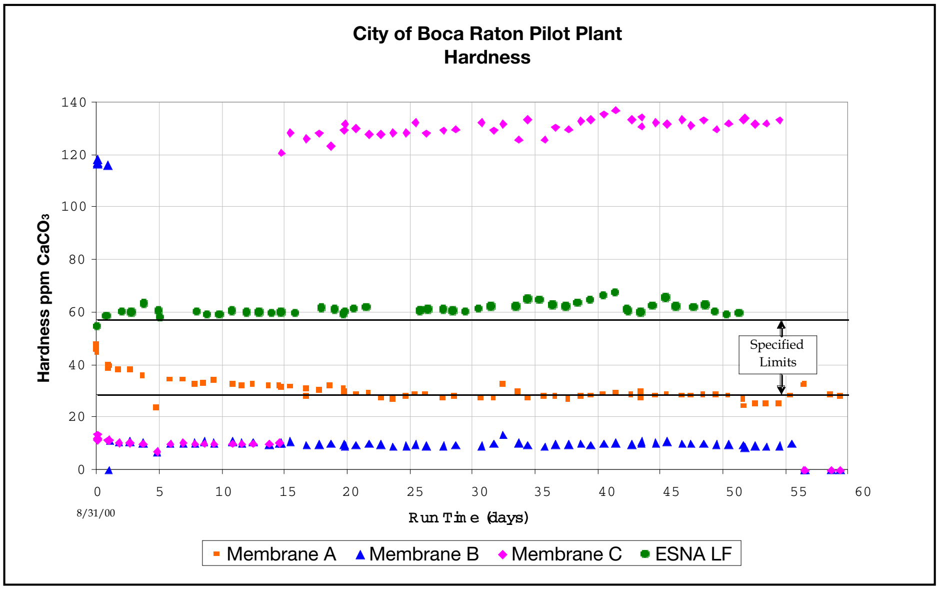

FIGURE 4. ESNA LF PERMEATE HARDNESS VS. TIME

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scalant, it was decided that membranes from three major membrane manufacturers should be tested side-by-side in the pilot test unit. The primary objectives of this testing were:

- to verify that the project water quality and performance;
- to confirm that membranes from approved membrane manufacturers could perform satisfactorily without acid and antiscalant;
- to confirm that membranes from approved membrane manufacturers perform in accordance with their performance projection programs.
- Permeate quality requirements for this side-by-side testing were based on the permeate quality objectives summarized in Table 1.

Three major nanofiltration membrane element manufacturers were asked to provide membrane performance projections for the full-scale membrane plant and the corresponding models of 4-inch diameter membrane elements for the side-by-side testing. Based on a review of these membrane performance projections, the maximum design transmembrane pressure for this project was set at 80 psi. In general, elements from all three manufacturers were able to operate within the required transmembrane pressure.

In contrast, only one of the membrane element manufacturers was able to produce permeate in the specified hardness range. A graph of permeate hardness versus time for the side-by-side testing is provided in Figure 4.

This graph illustrates that the membranes initially provided by both membrane manufacturers B and C produced permeate with hardness well below the specified range. Manufacturer C subsequently provided an alternate membrane element with lower hardness rejection.

These new membrane elements were installed in the pilot unit on day 15, and as shown in the graph, the permeate hardness produced by this membrane was well above the desirable range. Neither manufacturers B or C were able to offer hybrid designs acceptable to the city which could meet the permeate hardness target.

Only the membrane from manufacturer A produced water in the specified hardness range during the first round of side-by-side testing. Although the ESNA LF membrane was not originally included in the side-by-side testing due to price considerations, this membrane was subsequently added to the testing because of its more stable operating performance. As shown in Figure 4, although membrane A produced permeate hardness in the proper range, the membrane exhibited more susceptibility to humic acid fouling as evidenced by decreasing hardness passage with time. The ESNA LF membrane operated much more stably but produced permeate with a hardness just slightly above the target range.

2:1 Array Testing

While other membrane manufacturers were offered the opportunity to optimize the performance of their membranes to meet the required specifications, only Hydranautics indicated a willingness to make the necessary adjustments to its manufacturing process to optimize the performance of the ESNA LF to meet the permeate quality requirements. A series of ESNA LF membranes were produced with hardness rejections of 92, 95, and 98 percent. To demonstrate that the ESNA LF membrane could be made to meet the required permeate hardness range for this project, a 2:1 membrane array was tested with 95-percent hardness rejection ESNA LF membranes in the first stage and 92-percent hardness rejection membranes in the second stage.

The results of the 2:1 array testing are shown in Figure 5, which illustrates the first-stage, second-stage and composite total permeate hardness. The composite permeate hardness for the 2:1 array was approximately 60 milligrams per liter (mg/L), which is close to the middle of the desirable permeate hardness range.

Start-up Testing

The first unit to be placed into operation was Membrane Unit No. 1, which was started on August 11, 2004. Contractual Partial Utilization and the six-month performance testing schedule were initiated upon bacteriological clearance of Unit No. 1 five days later, with the first distribution of membrane-softened water to consumers. Start-up

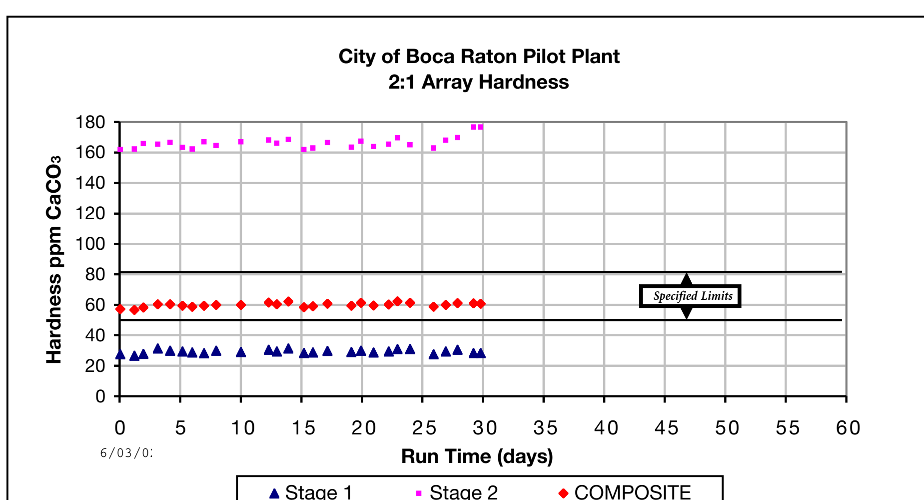

FIGURE 5. PERMEATE HARDNESS VS. TIME

and operation of Units 1, 2, and 3 proceeded in general accordance with the start-up schedule, with the permeate quality and membrane system performance meeting the specifications.

Membrane Unit No. 4 was started in November 2004. While it exhibited stable operation with no indication of fouling, the composite permeate hardness, which averaged approximately 100 to 110 mg/L as $CaCO_3$, exceeded the specified maximum of 80 mg/L. The membrane element manufacturer (MEM) acknowledged the noncompliance and advised that corrective action would be taken.

Unit No. 5 had been loaded with membranes from the same production run as Unit No. 4, so there was some concern that Unit 5 may exhibit similar out-of-spec performance. For this reason, Hydranautics requested that start-up of Unit No. 5 be delayed until corrective action could be taken and/or the membrane manufacturing procedures for these elements could be fine tuned; therefore, following start-up of Unit No. 4, the membrane elements in Unit No. 5 were preserved with a sodium bisulfite solution for this interim storage period. Hydranautics then made adjustments in the manufacturing and/or selection procedures for membrane elements for the remaining units.

Membrane Unit No. 6 was loaded and started in December 2004. Following startup, Unit No. 6 exhibited stable operation, with the permeate quality and membrane performance meeting the specifications. Start-up of Units No. 7, 8, and 9 followed, with all three units meeting water-quality and performance specifications.

During the last week of January 2005, an increase in transmembrane pressure was noted on several of the membrane units. The problem was found to be a breach in the cartridge filters, which allowed a small amount of sand to pass through to the membrane units. This sand was not apparent from silt density index (SDI) or turbidity readings.

The membrane units were shut down temporarily to allow repairs to the cartridge filters. Early detection of the sand problem limited the sand exposure to the lead elements in the units that were on line at the time.

Concurrently with the cartridge filter repairs, it was decided to remove the lead membrane elements in the exposed membrane units and consolidate them in Unit 1. The removed lead elements were replaced with non-lead elements from Membrane Unit 1.

The removed lead elements were cleaned by backflushing and vibration in a special apparatus provided by Poole & Kent and AEWT. Several elements were autopsied to confirm that the sand exposure was limited to the lead elements. There was no indication that any elements other than the lead elements had experienced any significant exposure to sand.

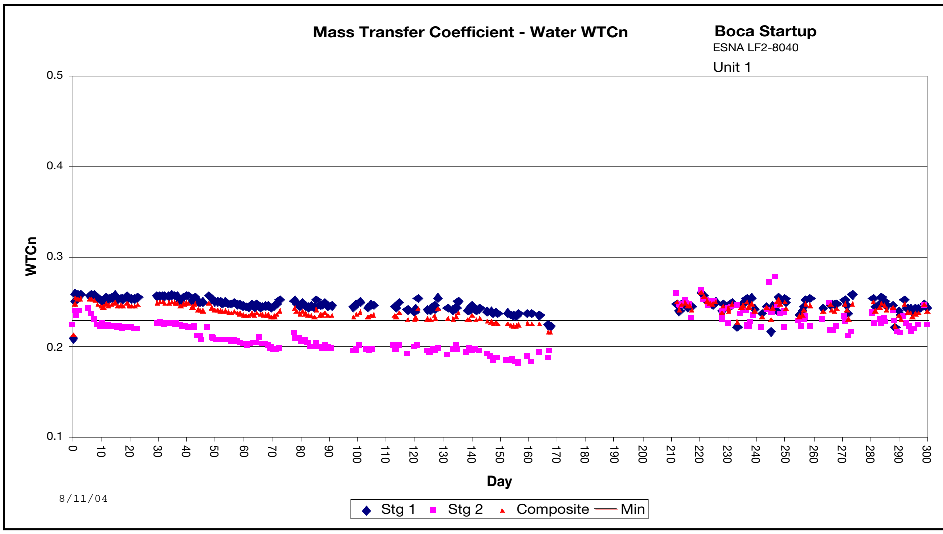

FIGURE 6. UNIT NO. 1 MASS TRANSFER COEFFICIENT VS TIME

The plant was gradually brought back on line during the third and fourth weeks of February. After the plant was restarted, no significant performance degradation was detected in any of the membrane units, including Unit No. 1, which contained all the lead elements from all exposed units. Performance on Unit No. 1 is discussed in more detail in the next section.

Membrane Unit No. 5 was started during the third week of February 2005. In comparison with Unit No. 4, Unit No. 5 start-up permeate quality and membrane performance were much closer to compliance with the specifications, operating at a transmembrane pressure (TMP) of approximately 67 psi and producing a composite permeate hardness of 81 mg/L.

Membrane Unit No. 10 was started up during the first week in March 2005, with the permeate quality and membrane performance meeting the specifications. Membrane Units No. 11 and 12 (the convertible units) were started up during the first week of April in the two-stage mode. Both units met permeate quality and system performance.

Performance Test Results

Operating Performance Results

Performance testing of the membrane units commenced on August 11, 2004, with the start-up of Unit No. 1. Performance data that were measured during the start-up, performance test, and initial commercial operation are shown in Figures 6 through 9 for Units 1 and 7. These data are generally representative of performance of the other units.

Start-up results for the full-scale plant confirmed that the membrane units could be operated satisfactorily without the use of either acid or anitscalant as pretreatment chemicals, as demonstrated in the pilot test program.

Figure 6 shows the mass transfer coefficient for the first-stage, second-stage, and overall composite performance for Unit No. 1 during the performance test period and the first 10 months of operation. A moderate downward trend in the mass transfer coefficient (approximately a 7 percent decrease) over the first 60 to 70 days of operation was noted and corrective action was initiated. A regimen of regular flushing and soaking of the membrane unit in permeate on shutdown was initiated around day 75. With this flushing program, the performance of the membrane system remained fairly stable over the next 90 days of operation, with only a 2.5-percent decrease in performance over this period. Operation of this unit and other units after the flushing program was initiated indicates very satisfactory performance without the addition of acid or antiscalant.

After 150 days of operation, the composite mass transfer coefficient had decreased by slightly over 10 percent since start-up, and membrane system cleaning was being considered. At that time, the sand incident described in the previous section was experienced. As discussed, membrane elements which were exposed to sand in other units were cleaned and then consolidated in Unit No. 1.

As shown in Figure 6, the mass transfer coefficient for Unit 1 actually improved as a result of the rearrangement with elements from other units. This graph also indicates that the performance of the membranes was not impaired by their brief exposure to sand.

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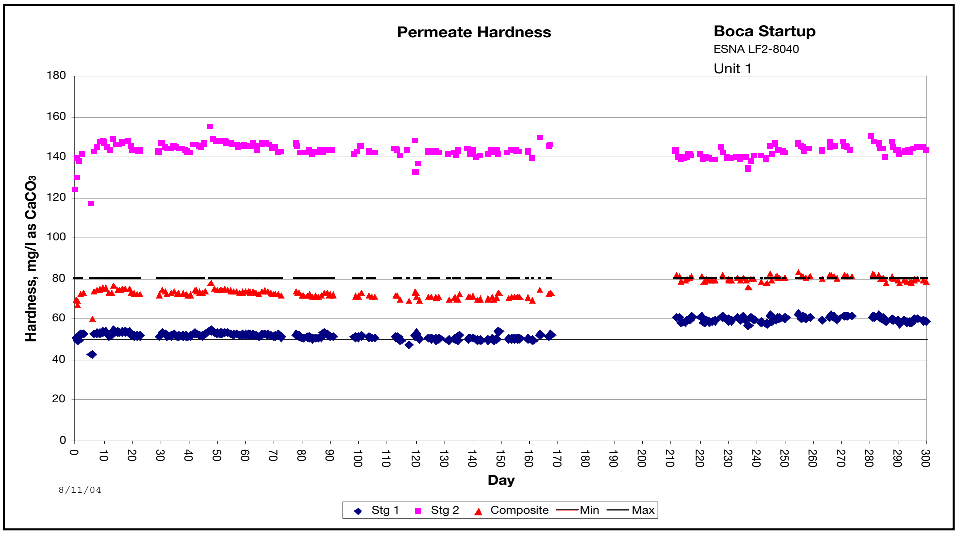

FIGURE 7. UNIT NO. 1 PERMEATE HARDNESS VS TIME

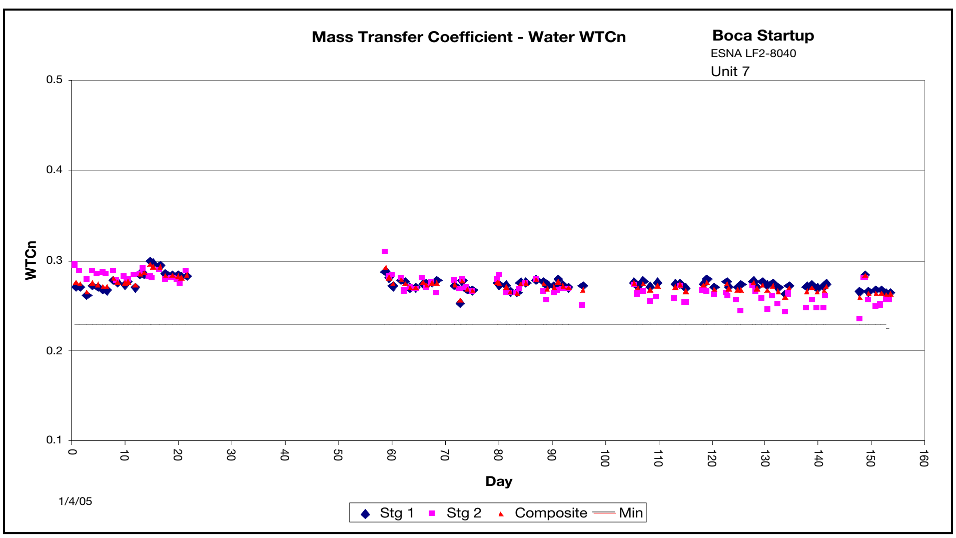

FIGURE 8. UNIT NO. 7 MASS TRANSFER COEFFICIENT VS TIME

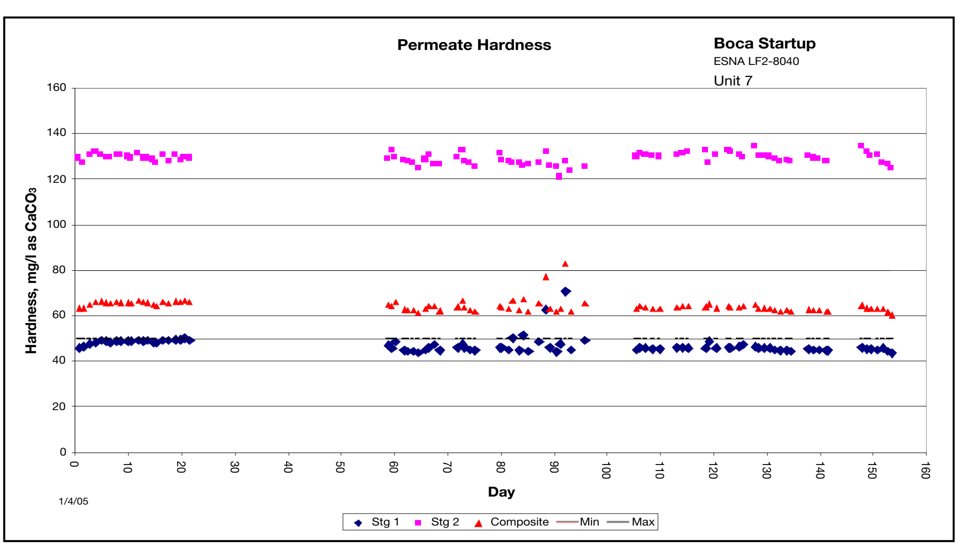

FIGURE 9. UNIT NO. 7 PERMEATE HARDNESS VS TIME

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As shown in Figure 6 and Table 2, the composite mass transfer coefficient of Unit 1 initially was in the range of 0.249 gfd/psi. As shown in Table 2, this translates to a required TMP of 73.3 psi, which is comfortably below the maximum specified TMP of 80 psi. This maximum specified TMP corresponds to a mass transfer coefficient of 0.23 psi/gfd and is shown as a solid reference line in Figure 6.

On the graph, if the mass transfer coefficient is above 0.23 gfd/psi, the TMP should be below the specified maximum. Figure 6 indicates that Unit No. 1 had a mass transfer coefficient (MTC) of 0.25 gfd/psi at start-up which gradually decreased to 0.23 gfd/psi over 120 to 160 days of operations, at which time cleaning was considered. The MTC increased back to 0.25 gfd/psi when the unit was restarted with cleaned membranes after the sand incident.

Analytical test results were correlated with conductivity measurements to produce the graph in Figure 7 of permeate hardness for Unit 1 over the same time period. The specified hardness window of 50 to 80 mg/L as $CaCO_3$ is indicated by two solid lines on the graph. Hardness measurements for the first and second stages were 50 and 145 mg/L, respectively, which resulted in a composite hardness of 73 mg/L. This is comfortably within the required range. After the membrane elements in the first stage were rearranged as a result of the sand incident, the permeate hardness increased to approximately 80 mg/L as a result of relocation of some lower rejection membranes from other units into Unit 1.

Graphs of mass transfer coefficient and hardness versus time for Unit 7 are presented in Figures 8 and 9 respectively. These results are fairly typical of the performance trends of all units. As shown in Figure 8, the mass transfer coefficient has remained relatively constant over the entire test period. The MTC at the end of the after 150 days of operation is within 5 percent of the initial MTC when the unit was restarted after the sand incident. Similarly, permeate hardness has remained very stable over the entire operating period, as shown in Figure 9.

Test results for all 10 of the primary 3.676 MGD treatment units are summarized in Table 2. These results indicate that the mass transfer coefficients for all units are above the target value of 0.231 gfd/psi with an average value of 0.278 gfd/psi. Similarly, the TMPs for all units were below the specified maximum value of 80 psi, with the average value being approximately 15 percent lower at 68.6 psi.

The average combined permeate hardness of all 10 of the primary units is 74.7 mg/L, which is within the specified hardness window of 50 to 80 mg/L as s. All the primary membrane units, with the exception of Unit 4, had permeate hardness values which substantially complied with the specified hardness window. The permeate hardness of Unit No. 5 is just marginally above the upper limit at 81.1 mg/L as $CaCO_3$. Unit No. 4 has a permeate hardness of over 95 mg/L as $CaCO_3$. Hydranautics has agreed to take corrective action to bring the total permeate hardness of Unit No. 4 into the required hardness window between 50 and 80 mg/L.

	Mass Transfer Coef. (MTCw)			TMP	Permeate Hardness (mg/L as $CaCO_3$)		
	Stage 1	Stage 2	Composite	(psi)	Stage 1	Stage 2	Composite
Unit 1	0.256	0.227	0.249	73.3	52.4	143.3	73.1
Unit 2	0.262	0.229	0.255	71.3	54.8	155.8	76.2
Unit 3	0.313	0.291	0.308	61.0	56.6	141.7	75.1
Unit 4	0.280	0.293	0.283	66.2	70.7	176.8	95.9
Unit 5	0.280	0.252	0.274	67.4	54.1	166.2	81.1
Unit 6	0.335	0.290	0.326	65.9	57.5	143.3	75.9
Unit 7	0.277	0.279	0.277	67.2	47.6	129.4	65.0
Unit 8	0.281	0.278	0.281	69.0	49.7	140.3	68.7
Unit 9	0.257	0.268	0.259	72.7	47.4	138.9	67.7
Unit 10	0.276	0.203	0.263	72.7	51.1	144.7	68.1
Average:			**0.278**	**68.6**			**74.7**
Spec:			>0.23	<80			50 - 80

TABLE 2: CITY OF BOCA RATON MEMBRANE UNIT START-UP PERFORMANCE

Analytical Test Results

Analytical test results for the primary membrane units are summarized in Table 3. The primary parameters of interest are hardness, color, and THM and HAA formation potential.

Hardness testing was performed by an outside laboratory, as well as the city's state-certified laboratory. Analytical test results for hardness were correlated with conductivity readings, and permeate hardness plots versus time, similar to the one shown in Figure 4, were developed for each membrane unit.

The data in Table 2 summarize daily hardness values calculated from permeate conductivity readings, averaged over the 30-day performance test for each membrane unit. The hardness data in Table 3 is the average of all hardness values for each sampling event for each unit, as reported by the outside laboratory, averaged with the analyses provided by the city for each membrane unit.

Although on a case-by-case basis there are some variations between the hardness values in Table 2 and Table 3, the average hardness values for all membrane units is 75.4 mg/L in Table 3, as compared to 74.7 mg/L by the conductivity method (Table 2), which are in relatively good agreement. The hardness data in Table 3 indicates that eight of the 10 primary membrane units produced permeate with hardness in the specified 50-to-80-mg/L window.

Although the reported hardness results for Unit 5 in Table 3 are somewhat higher, hardness results reported by the city and calculated by the conductivity method in Table 2 indicate that the hardness is only slightly over the 80 mg/L, at 81 to 82 mg/L, so this unit is close to compliance. Only Unit No. 4 has permeate hardness which exceeds the 80 mg/l maximum limit by an appreciable amount.

Another parameter of primary importance to the city is color. With the lime softening process, the city typically produces water with color in the 12 to 15 CU range, which is marginally below the state secondary standard for color of 15 CU. Since color analysis from the outside lab had only a minimum detection limit of 5 CU for color, more accurate test results from the city's analytical lab were used in Table 3.

These results indicate that permeate color was very low, averaging just over 1 CU for all units. All the membrane units except for Unit 3 had reported color values of 1 CU or less. Since color readings are primarily associated with dissolved organics, the fact that the TOC value of 0.9 mg/L is less than 1.0 mg/L similar to all the other membrane units, it would be anticipated that Unit 3 should have a similar color reading as the other units, and that there may be some reporting error for color for Unit No. 3.

Results of THMFP and HAAFP, as reported by the outside laboratory and averaged with the city's analytical results, are summarized in Table 3. The THMFP values for all units were well within the required maximum concentration of 42 micrograms/L. The HAAFP values in Table 3 are also below the maximum specified limit of 30 micrograms/L.

Based on these results and verified by the city, it was concluded that all the primary units met the quality requirements for hardness, color, THMFP, and HAAFP with the exception of Unit No. 4, which did not meet the required permeate hardness. Hydranautics has agreed to take corrective action on Unit No. 4 to bring the permeate hardness into compliance.

Pretreatment Cost Savings

One of the primary design objectives for the project was to reduce or eliminate the need for acid pretreatment due to the costs, operational complications, and potential hazards associated with handling the large amounts of acid necessary for a plant of this size. A great deal of effort was expended to develop this concept, demonstrate the full-scale feasibility, and optimize membrane performance with respect to the city's water-quality goals on the high-fouling, potential raw-water supply.

These efforts included acrylamide gel electrophoresis investigations in the laboratory conducted by Dr. Harvey Winters on various antiscalants and membranes using the city's raw water, as well as over three years of pilot testing during the design and construction phases. Ultimately it was found that operation was actually more stable without any chemical pretreatment than with any of the chemical pretreatment protocols that were evaluated, so the plant was started up and is now operating successfully with no chemical pretreatment. This finding has substantial cost and operating benefits that are being realized in the full-scale operation.

The original basis of design during the preliminary design called for feedwater pH adjustment from the raw-water pH of 7.2 to a feed pH of 6.2, which is typical of nanofiltration plants treating Biscayne Aquifer raw water in the South Florida area. This would require a dosage rate of approximately 110 mg/L of sulfuric acid.

At the city's total average daily demand (ADD) of 45 MGD and the design blend rate of 2:1 permeate:lime softened water, the total permeate production would be 30 MGD, which would require a feed flow of approximately 35.3 MGD. This would result in an average acid usage rate of approximately 34,820 pounds per day of 93 percent sulfuric acid. Based on a unit price of $0.045 per pound, this translates to an annual cost of approximately $572,000.

Besides the acid-related cost savings, there is also a substantial cost savings from the reduced need for caustic for post-treatment pH adjustment. Post-treatment pH adjustment consistent with the above-described pretreatment pH adjustment, to a

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final permeate pH of 8.3, assuming a unit price of $0.15 per pound for 50 percent sodium hydroxide, would result in an annual cost increase of approximately $97,000 over the current caustic usage rates at the plant.

There is also a chemical cost savings from eliminating pretreatment antiscalant. The original preliminary design basis called for an antiscalant pretreatment dosage of 3 mg/L. Assuming a unit price of $0.80 per pound, this would result in an annual cost of approximately $258,000.

The total direct cost savings from the elimination of chemical pretreatment for this facility can be estimated conservatively at approximately $927,000 per year, which does not consider indirect costs associated with operation and maintenance of the chemical storage and feed systems, training, cleaning, etc. Also, a substantial day-to-day safety risk source to the operations staff has been completely avoided at the facility.

Cost Savings Associated With Reduced Operating Pressure

As shown in Table 2, noted in Section 4.1, the specified maximum TMP for the full-scale units was 80 psi. The actual average TMP demonstrated during performance testing was approximately 69 psi. Under the ADD condition feed-flow rate of approximately 35.3 MGD, and assuming a unit power cost of $0.045 per kilowatt-hour, this feed-pressure differential results in an annual savings of approximately $48,000 per year.

Conclusion

The Glades Road Water Treatment Plant is now the largest nanofiltration facility in successful operation in the world. Numerous challenges were encountered along the way, but through a comprehensive program of pilot testing, design development, and construction supervision, these challenges were overcome and an optimum membrane treatment solution was developed.

Initial severe colloidal fouling problems were solved through a combination of wellfield rehabilitation and installation of multimedia filtration. When traditional chemical pretreatment options failed due to unacceptably high fouling rates associated with naturally occurring organics, an unconventional approached was adopted that eliminated pretreatment chemicals and put the naturally occurring organics to work as scale-controlling agents. Not only does this approach simplify operation and avoid potential problems associated with acid storage and handling, but more important as shown in the previous section, it significantly reduces operating costs.

Other significant factors which contributed to the success of this project were the introduction of low-fouling nanofiltration membranes and the optimization of membrane performance to meet the city's treatment objectives. These new low-fouling ESNA LF nanofiltration membranes were developed specifically for this project and were the first membranes used to demonstrate the viability of long-term operation without acid and antiscalant pretreatment. In addition, the performance of the membranes was optimized to not only meet the city's water-quality goals, but the lower operating pressure of these membranes as compared to the specified values resulted in savings in operating power costs.

The extended start-up and performance testing program specified for the project allowed the membrane manufacturer to optimize the performance of the membranes based on full-scale operating data. Stable operation of these membranes during the first year appears to indicate that this success will be long-term.

In addition to achieving all the water-quality and capacity goals established by the city at the onset of the project, the innovations discussed above will result in annual cost savings for the city of approximately $1 million dollars per year.