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Operation Optimization and Membrane Performance of a Hybrid NF Plant

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Abstract

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Low fouling nanofiltration membranes are being used at a 10 mgd plant in Pompano Beach, FL to meet difficult treatment objectives for a high fouling, colored feedwater. Based on extensive pilot testing and process design considerations, the hybrid design NF system was selected and installed at site. In summer 2009, the plant was re-commissioned and was able to meet the required feed pressure and permeate quality targets during the 60 day Performance Acceptance Test for the 5 trains. The challenging permeate quality targets required iron to be reduced below 0.2 ppm and disinfection by-product formation potential to very low levels, while keeping calcium between 8-33 mg/l and alkalinity between 20-61 mg/l as $CaCO_3$. This type of partial softening requires an optimized form of NF membrane to meet all the requirements.

The plant has now been operational for more than one year. There have been some indications of slow, but steady fouling of the membrane surface. For example, in one train the trans-membrane pressure rose from 75 psi to 90 psi in about 10 months. During this period there was very little change in differential pressure, but salt passage was decreasing. Early in the commissioning process, fouling was minimized by reducing pH and reducing the antiscalant dosing. Most recently, the plant has used a high pH cleaning, which dropped pressure back below 90 psi. Analysis of the foulant suggests the presence of colloidal iron and aluminum silicates, as well as organic and phosphate based colloidal material. SEM analysis indicates that the foulant is heavy enough to cover the surface of the membrane, but not plug the spacer material. A detailed analysis of the foulant and cleaning effectiveness is reviewed.

The plant continues to operate well and meet the City's water treatment objectives. Further analysis is being made of the fouling to better restore membrane performance and prevent further fouling trends.

1. Introduction

Membrane technology has become a valuable advanced water treatment process to purify difficult water sources for potable use. The source waters can vary from highly saline seawater to low salinity ground or well waters. In the latter case, treatment objectives often involve the selective removal of certain contaminants such as hardness, color, iron and disinfection byproduct precursors to meet drinking water standards and improve water quality aesthetics. In this case, nanofiltration membranes are the product of choice to treat the water, but a detailed understanding of the membrane and water source is needed to ensure a successful application of NF technology.

One plant that dealt with this difficult requirement is the Pompano Beach Nanofiltration NF plant in Florida. The original plant was constructed and started operation in 2002 using a hybrid design of ultra-low pressure RO and NF membranes. However, due to operational and design issues the feed pressures were relatively high and rejection of feedwater constituents was not meeting desired targets (Kiefer 2009). A retrofit was made to the Pompano Beach Water Treatment plant to ensure that it could meet very strict treated water quality targets, while producing 10 mgd of treated water at lower energy consumption. The goal of this project was to treat the Biscayne Aquifer with TDS around 500 mg/l, to meet strict requirements on bicarbonate (25-75 mg/l), calcium (8-33 ppm), color (<3 CU), iron (<0.2 mg/l), TDS (<250 mg/l), Total THM-FP (< 40 ug/l) and Total HAA-FP (<30 ug/l). To achieve this goal, a hybrid membrane design was selected, using Hydranautic's low fouling ESNA1-LF and ESNA1-LF2 elements. The type and combination of two different NF elements were selected to maintain calcium and alkalinity concentrations above the specified minimum values, while not exceeding the specified concentration for iron, color and disinfection byproduct precursors. (Kiefer 2009)

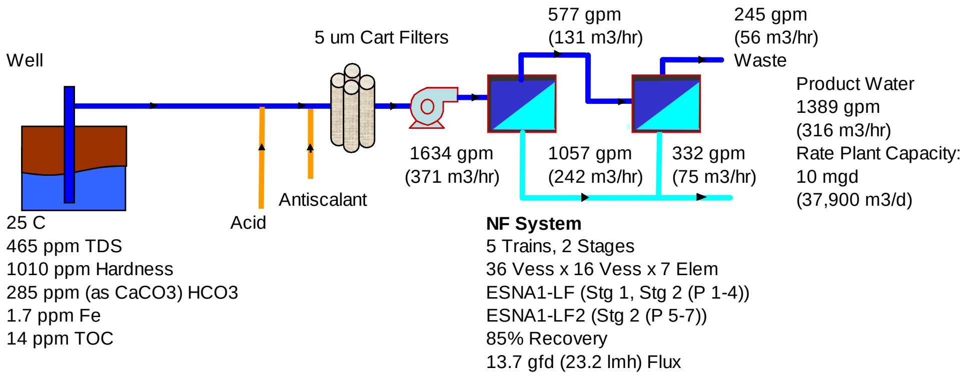
The initial operation of this system led to some practical issues regarding membrane performance and operating costs. In the initial design of the process and pilot testing, it was planned to acidify to pH 6.2 to 6.5. However, upon execution of the full-scale plant, the cost of sulfuric acid went up significantly, making the original pH design operational cost much more expensive. As a result, the plant opted to reduce the usage of sulfuric acid by pH adjusting from 7.2 to only 6.8. There are other plants on the same aquifer that have operated successfully without acid (e.g. Boca Raton & Plantation). However, other facilities in the area have had varying levels of success in their ability to eliminate or reduce their use of acid. At this facility it quickly became apparent that the rate of fouling increased faster than desired and corrective action would be required. This paper reviews the optimization work that was done for the plant.

2. Pompano Beach NF Plant Design

The Pompano Beach NF Plant has recently been upgraded to improve the performance and operation of the plant. (Kiefer 2009) The system, as it is currently configured, started operation in March 2009. The system consists of five trains with each train arranged in a two-stage array of 36x16-7M vessels (Figure 1). The 1st stage uses ESNA1-LF nanofilter elements. The 2nd stage uses ESNA1-LF nanofilter elements in the 1st four positions and ESNA1-LF2 elements in the last three positions. The LF2 membrane is a "looser" membrane which allows for reduced feed pressure and allows for more calcium and bicarbonate to pass through to the permeate, and the hybrid still has sufficiently good rejection of iron to < 0.2 ppm in the permeate. The process flow scheme for the plant is shown in Figure 2.

Figure 1 Pompano Beach 10 MGD NF Plant

Figure 2 Design Process Scheme for the Pompano Beach NF Plant



The challenge for this plant is the presence of high levels of hardness, iron, color and total organic carbon in the feedwater (Figure 3). The organic matter gives rise to high levels of disinfection by-products when the water is chlorinated in the distribution system and can rapidly foul many RO membranes, while causing aesthetic issues for consumers. The permeate quality required for this plant was also difficult to achieve. The actual feed, concentrate and permeate constituents and the permeate water specified targets are shown in Table 1. One of the biggest challenges with this feedwater composition and product specification is the need to have high rejection of iron and organics, while having moderate rejection of the hardness and alkalinity. Various papers have documented the successful use of new low fouling NF membranes to accomplish both objectives. These membranes can be tailored to have hardness rejection in the range required for these applications. (Bartels 2008, Kiefer 2004, Kiefer 2006)

Figure 3 Highly Colored Feedwater of the Pompano Beach Plant

Table 1 Actual Water Quality and Specified Permeate Quality

Item	Feed	Concentrate	Permeate	Specification
Calcium (mg/l)	96	763	9.7	8 – 33
Magnesium (mg/l)	5	42	0.5	
Sodium (mg/l)	34	98	18	
Iron (mg/l)	1.7	10.3	0.15	< 0.2
Alkalinity (mg/l as CaCO3)	154	1090	36	20 - 61
Sulfate (mg/l)^	110	510	12	
Chloride (mg/l)	45	301	45	
Color (CU)	120	650	<0.5	< 3
Hardness (mg/l as CaCO3)	262	2080	26	
TDS (mg/l)	465	3420	67	< 250
TOC (mg/l)	14	120	0.4	< 1
Trihalomethane FP (mg/l)			<0.01	< 0.040
Haloacetic Acid FP (mg/l)			<0.001	< 0.030
TMP (psi)*	87			< 90

^ Sulfate after acid addition

* TMP Trans-Membrane Pressure (feed – permeate) at the end of the 60 day acceptance test.

3. **Plant Operation and Performance Trends**

The plant was successfully started with an average TMP of 80 psi (5.5 bar) for all five trains at day 1 and 87 psi (6.0 bar) at day 60. The maximum target TMP was 90 psi (6.2 bar). Table 1 shows that the actual permeate quality met and exceeded the requirements of the specification by the selection of four higher rejecting NF membranes followed by three lower rejecting NF in each pressure vessel of the 2nd stage. The use of all ESNA1 LF membranes would have lowered calcium and alkalinity below target values. The minimum permeate calcium level was met at 9.7 ppm as Ca and the target was 8 to 33 mg/l. The minimum permeate alkalinity level was met at 36 mg/l as CaCO3 and the target was 20 to 61 mg/l. While achieving the minimum calcium and alkalinity levels, the 0.15 mg/l of permeate iron met the target of < 0.20 mg/l.

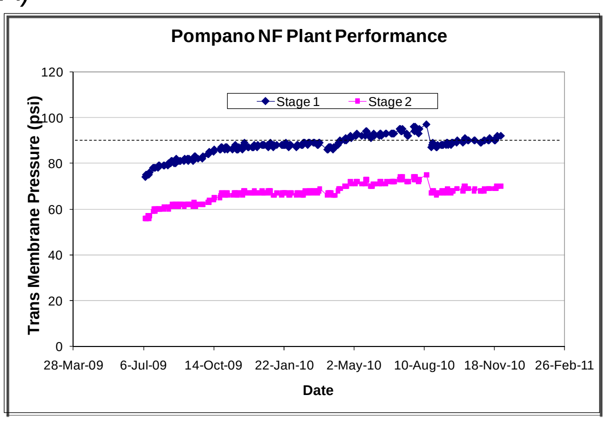
Graphs of the normalized performance and TMP of Train # 4 for the first 16 months of operations are shown in Figure 4. It can be seen that there has been a slow but steady fouling of the membranes in both stages during the first sixteen months of operation. After 10 months of operation, the TMP rose to the target value of 90 psi (6.2 bar), but it was decided to hold off on cleaning to control operating costs while being cautious not to adversely impact long term operations.

After 13 months of operation (August 2010), the TMP increased to 98 psi (6.8 bar) (Figure 4A) and at this point, the decision was made to perform the 1st cleaning of this train based on reducing TMP to reduce energy cost, increasing % salt passage so as to increase the calcium passage above the minimum target level, and hoping to target cleanings for only once a year. Only a high pH cleaning of both stages was performed to reduce TMP and feed pressure. As a result of this cleaning, the TMP dropped back to 87 psi (6.0 bar) and the normalized 1st stage permeate flow was restored to 84% of base-line value and the 2nd stage was restored to 80% of baseline value. This pressure is well below the maximum end-of-useful life feed pressure of 140 psi (9.7 bar) (where 22 psi (1.5 bar) was used as permeate back-pressure), previously experienced at this plant when the higher rejection membranes were used. Likewise, the normalized % salt passage (Figure 4C) also dropped 34% for the 1st stage and 45% for the 2nd stage in 13 months before cleaning. As a result of the cleaning, the first stage normalized % salt passage was restored to 84% of baseline and the 2nd stage was restored to 81% of the baseline. The cleaning, as expected, caused an increase in normalized salt passage, suggesting that the foulant is the major cause of the increased rejection and is evenly distributed from the front to the back of the system. It should be noted that it is normal for rejection to improve over time on these organic-laden Biscayne aquifer feed waters. The normalized differential pressure has increased slowly but at a steady rate of only 4 psid (0.3 bar) for 13 months and the cleaning reduced the differential pressure by a few psid.

After 16 months of operation (November 2010) and only one cleaning after 13 months, the 1st stage TMP was 91 psi (6.3 bar) (Figure 4A). The 1st stage normalized permeate flow had dropped 21% from start-up and 25% for the 2nd stage (Figure 4B). The normalized % salt passage had essentially stabilized back to the levels observed prior to the cleaning 3 months earlier at 7% for the 1st stage and 4.5% for the 2nd stage (Figure 4C). The normalized delta P for both stages had increased only 4 psid (0.3 bar) per stage, indicating there is a low level fouling of the feed spacer, but very much under control (Figure 4D).

Figure 4 Pompano Beach NF Plant Performance (A) Trans Membrane Pressure, (B) Normalized Permeate Flow, (C) Normalized Salt Passage, and (D) Normalized Differential Pressure.

A)



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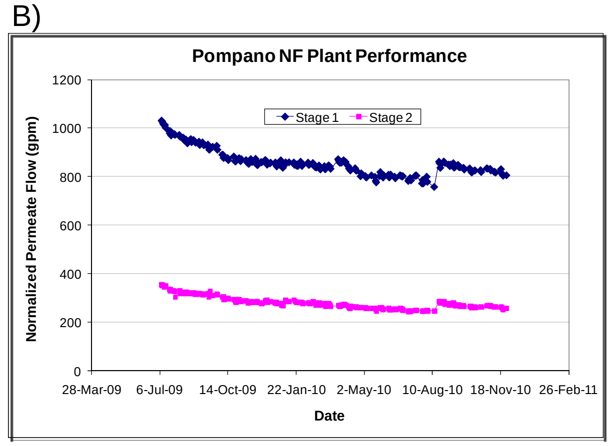
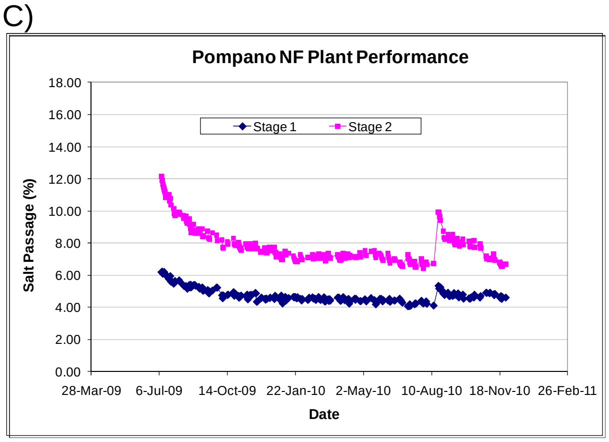


C)



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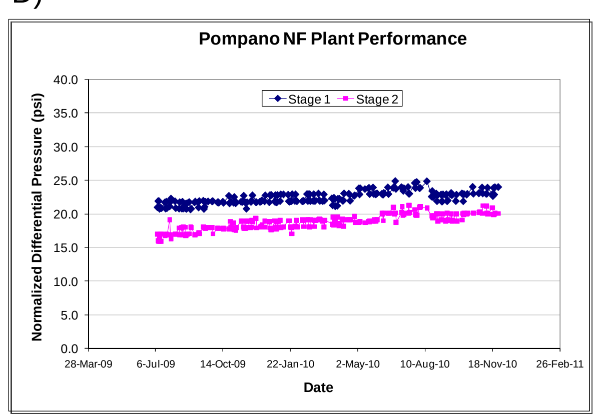
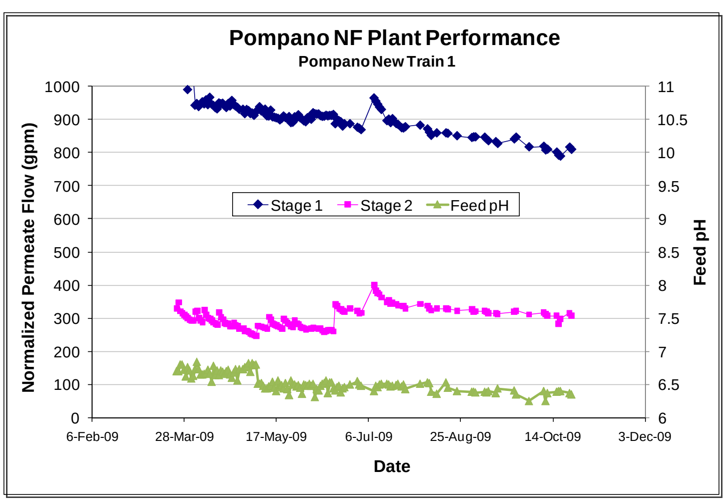


4. **Fouling Issues and Analysis**

At the start of the Performance Acceptance Test (PAT), fouling was observed in the first train started that had not shown itself during the earlier pilot testing. In the first weeks of the PAT of Train 1, the normalized permeate flow started to decline rapidly. (Figure 5) Past experience at near-by plants indicated there may be an issue with antiscalant dosing and operations at higher pH. It was felt that this may be due to a "gel layer" foulant being formed on the surface of the membrane. This "gel layer" acts as a dynamic membrane layer on top of the RO/NF membrane, and as it develops over time, it increases the feed pressure requirement and decreases the passage of salt through the membrane. This "gel layer" may be comprised in part of naturally occurring organics (NOM) in the Biscayne Aquifer and the man-made organic antiscalant. Empirical evidence from other water plant operations indicates that NOM also have antiscalant properties, which allows for the reduction of the antiscalant dosage which was instituted here. (Bartels 2008) Therefore, it was felt that the amount of antiscalant could be reduced slightly, without inducing rapid scaling at the end of the 2nd stage.

The initial dosage of antiscalant for Train 1 at start-up was 2.1 ppm and the feed pH was 6.8. This is slightly different than the pilot condition, which showed stable performance at pH 6.2. One of the reasons for this increase in pH was the 350% increase in the cost of sulfuric acid. Other plants have also been trying to reduce the amount of acid added to the feedwater (Lai 2009). After 43 days of continuous operation, the normalized permeate flow declined approximately 12% and 25% for stage 1 and stage 2, respectively (Figure 5). At the same time, the salt passage also declined 7% and 12% respectively for stages 1 and 2. Controlling the rate of development of this "gel layer" became important in controlling the rate of fouling. Corrective action was taken on day 43 of Train 1 operation to reduce the rate of fouling. The feed pH was decreased from 6.8 to 6.5 and the antiscalant dosing rate decreased from 2.1 to 1.5 ppm. This resulted in a significant reduction in the rate of fouling, particularly in the 2nd stage, as evidenced by improved normalized permeate flows (Figure 5). The antiscalant dosage was lowered a second time on June 3, 2009 from 1.5 to 1.0 ppm with a smaller but noticeable decrease in the rate of fouling.

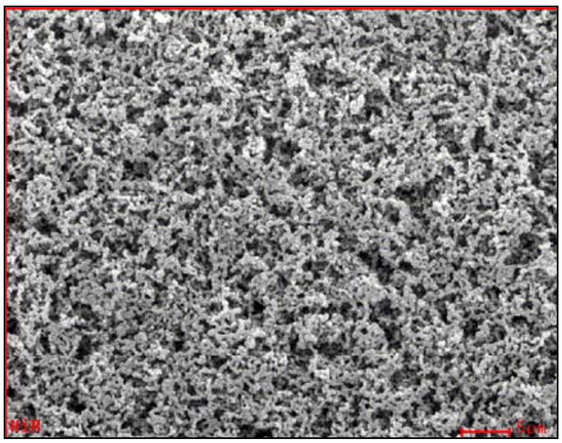
Figure 5 Initial Operation: Normalized Flow for Train 1



To evaluate the make-up of the colloidal foulants in the NF feed, a large quantity of feedwater was run for 3.5 hours through a 0.45-micron SDI pad, which was then analyzed by SEM (scanning electron microscope) and EDAX (energy dispersive x-ray). Much of the foulant collected on the 0.45 micron SDI pad (Figure 6B) appears to be colloidal in nature and is a mixture of inorganic and organic matter. Given the large volume of water required to collect this sample and that SDI's on site are very good and range from 1.4 to 2.2, it is theorized that colloidal fouling is not a major contributor to the fouling rate, and if so, is limited to the lead element of the 1st stage. The primary inorganic constituents measured by EDAX were mainly calcium, phosphorous, iron, aluminum, and silica but it is important to note these items seem not to be a system problem. The origin of phosphorous could be from the feedwater or possibly the phosphonate-based antiscalant and exists in the form as a calcium phosphate scale. The iron, aluminum and silica on the SDI pad would probably be attributed to being naturally occurring components of clay. Iron is consistent with the orange color on the element (Figure 7A).

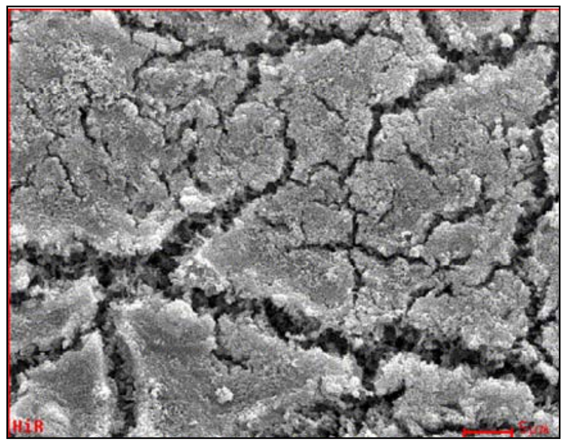
Figure 6 SEM and EDAX Analysis of Foulant on SDI Pads (3.5 hrs of filtration)

A) SEM photo of a new SDI pad



Element	Wt%	At%
CK	42.25	50.19
NK	12.31	12.55
OK	39.04	34.82
NaK	00.92	00.57
ClK	01.48	00.60
KK	00.41	00.15
CaK	02.02	00.72
FeK	01.55	00.40
Matrix	Correction	ZAF

B) SDI Pad after 3.5 hrs of filtration



Element	Wt%	At%
CK	33.43	46.99
NK	07.32	08.83
OK	29.92	31.57
NaK	00.75	00.55
MgK	00.34	00.23
AlK	01.23	00.77
SiK	02.03	01.22
PK	05.15	02.81
ClK	00.49	00.23
KK	00.21	00.09
CaK	07.70	03.24
FeK	11.42	03.45
Matrix	Correction	ZAF

4.1 Membrane Analysis

Two elements were removed from Train 1 30 days after start-up, one a lead element from stage 1, and a tail element from stage 2. The retest results are shown in Table 2. The lead element, A1413271, was retested at standard test conditions and showed a 21% decrease in flow and a 44% increase in rejection when compared to original factory data. Tail element A1546687 was also retested and showed a 58% decrease in flow and a 38% decrease in rejection when compared to original factory data.

Table 2 Factory Testing at Standard Test Conditions of Returned Lead and Tail Elements

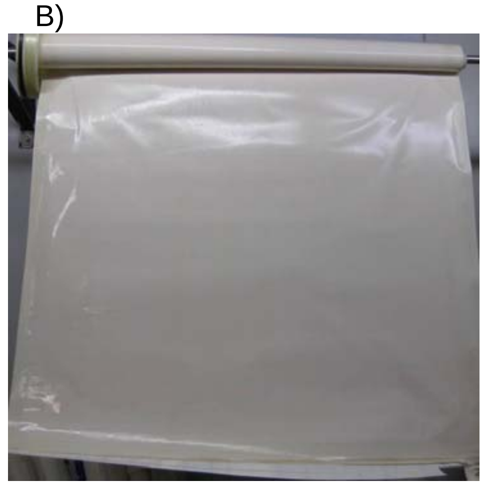
Standard Re-test Performance Data							
SERIAL NUMBER	ORIGINAL		RE-TEST			% CHANGE	
	Rejection	Flow (GPD)	Rejection	Flow (GPD)	dP (psi)	Salt Passage	Flow (GPD)
A1413271	87.30%	10,011	92.89%	7,928	4.0	-44%	-21%
A1546687	84.90%	12,791	79.22%	5,380	3.9	+38%	-58%

Upon inspection, the tail element was lightly discolored on the outside with a brown foulant (Figure 7A). The tail element was cut open and the membrane sheets appeared relatively clean, with only a fine layer of foulant (Figure 7B).

Figure 7 Returned Tail Element (A) Outside and (B) Membrane Leaf

A)

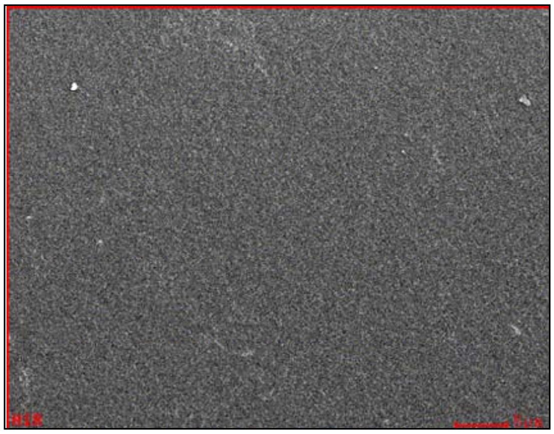
B)



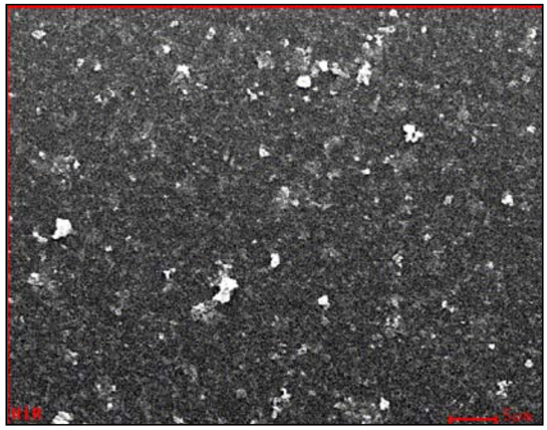
Samples of the membrane were removed and analyzed by SEM and EDAX surface analysis. The SEM micrographs of the fouled surface (Figure 8B) show that the foulant covers the membrane surface completely. The composition of the foulant was determined by EDAX, as shown in the charts of Figure 8. The main difference between the fouled membrane and the unused membrane is the greater amount of carbon and some iron and aluminum. The carbon is due primarily to the organic foulant, such as the NOM, or possibly the organic polymeric antiscalant. The iron and aluminum is likely due to the presence of colloidal material such as iron hydroxide or aluminum silicate. The calcium could be the result of calcium phosphate or calcium carbonate. There is no indication that the colloidal material or calcium scale is a major contributor to the system fouling. The organic matter is a definite factor in the fouling as corroborated by the effectiveness of caustic cleaning.

Figure 8 SEM/ EDAX Analysis of the Tail Membrane

A) SEM - Fresh Unused Membrane



B) SEM - Fouled Membrane from Site



Element	Wt%	At%
CK	57.75	69.14
NK	05.78	05.93
OK	16.55	14.87
NaK	06.41	04.01
SK	13.18	05.91
ClK	00.33	00.13
Matrix	Correction	ZAF

Element	Wt%	At%
CK	70.11	79.25
NK	06.05	05.87
OK	11.76	09.98
AlK	00.14	00.07
SK	10.08	04.27
ClK	00.15	00.06
CaK	00.93	00.32
FeK	00.76	00.19
Matrix	Correction	ZAF

Cell testing was also conducted on membrane samples extracted from the tail element A1546687 as the tail element should exhibit the highest level of “gel layer” foulant and had the highest drop in flow at the factory retesting. Cell testing confirmed the membrane flux was in fact low. A thirty-minute pure water flush at 2 microsiemens-cm and pH of 4.6, followed by a 2 hour soak in pure water, was conducted on the membrane samples. Flux increased approximately 88% and rejection increased 43% after the permeate flush/soak. The increase in rejection can be attributed to the increase in flux which further dilutes the salts which pass through the membrane. This improvement in flow and rejection are also confirmed at the site when the trains are allowed to soak in permeate.

5. **Cleaning**

During the first 13 months of operation, there were no cleaning events for Train 4. After evaluation of the foulant and the experience with the performance of the system, it was determined that a caustic cleaning would be done in order to lower the TMP of 98 psi (6.8 bar) and also beneficially increase the passage of hardness to levels higher than the target 8 mg/l as Ca. This is also a result of the foulant analysis performed earlier, which showed a high level of carbon-based organic matter on the membrane surface. The cleaning was done with a high pH solution of NaOH adjusted to pH 11.3 and heated to 32° C. The operators alternated between recirculation and soaking during the day shift. They then left the train overnight to soak with chemical while heated. In the morning, the cleaning solution was recirculated again. The train was put back on line after thorough permeate flushing.

The positive results on operation can be seen in the normalized data graphs of Figure 4. As a result of this cleaning, the TMP dropped back to 87 psi and the normalized 1st stage permeate flow was partially restored to 84% of base-line and the 2nd stage to 80% of baseline. (Table 3) In this case, a low pH cleaning with HCl acid at pH of 2.5-2.7 was not performed as the usefulness of the low pH cleaning was still being evaluated. Current thought is to follow up with a low pH cleaning only once every two or three. However, as is common on these feed waters and when the membrane is not completely cleaned of the "gel-layer" foulant, the rate of fouling after the clean is initially rapid, but then stabilizes again. From these results, it is possible that a more aggressive cleaning may be considered or the feed pH can be lowered to reduce the rate of "gel-layer" build-up to stabilize performance if the rate of TMP increase over time is higher than desired.

Table 3 Pompano Normalized Cleaning Results for Train 4

	Salt Pass Initial	Salt Pass 13 mos	Salt Pass Change	Salt Pass Post Clean	Salt Pass Recovery
	%	%	%	%	%
Stage 1	6.2	4.1	-34%	5.3	85%
Stage 2	12.2	6.7	-45%	9.9	81%

	Norm Flow Initial	Norm Flow 13 mos	Norm Flow Change	Norm Flow Post Clean	Norm Flow Recovery
	gpm	gpm	%	gpm	%
Stage 1	1030	757	-27%	861	84%
Stage 2	356	247	-31%	285	80%

6. **Conclusions**

It can be seen that the hybrid NF membrane design was able to meet the project requirements to treat this difficult, highly colored feedwater. Although there was some initial fouling, the TMP was still below the 90 psi mark after the first 60 days of operation and pressure stabilized reasonably well after that. The low pressure NF membranes were able to produce permeate which comfortably met the requirements for organic removal, low color and low disinfection by-product formation potential. The hybrid membrane design challenge of "threading the needle" was successfully met by allowing sufficient hardness and alkalinity passage to meet the minimum targets and still keep iron passage less than the maximum 0.2 mg/l.

Operational data showed that there was some rapid fouling initially, which was mostly due to the operation at a higher feed pH. To resolve this, the sulfuric acid dosing was increased to lower feed pH, while the antiscalant dosing was reduced but kept at a safe level. This remedial action helped stabilize the performance. The final dosing rate of acid and antiscalant was a trade-off between lower chemical costs and more frequent membrane cleaning.

Generally, the trains were able to run for over a year without cleaning. The fouling that has occurred during the bulk of the operation appears to be equally dispersed through the NF system. Both stage 1 and stage 2 have shown about the same decrease in normalized flow. When a high pH clean was performed, critical operating parameters were restored to 80-85% of baseline conditions. A more aggressive cleaning will be needed to further restore performance.

Finally, it can be seen that the full-scale plant is performing similar to the pilot plant. The operation at a slightly higher pH (6.5 actual versus 6.2 in the pilot) seems to be causing slightly higher rates of fouling. This is considered tolerable, since the higher pH operation is resulting in significant chemical savings. When compared to the operation of the original plant, this retrofit is providing significant cost savings. As previously stated, (Kiefer 2009) the City is saving about $106,700 per year in operating costs (at a power cost at $0.10/kwhr) due to the use of the new low pressure, low fouling NF membranes.

7. References

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