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Cost and performance of a hollow-fiber loose nanofiltration membrane pilot-plant for aerated groundwater treatment

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ABSTRACT

The cost and performance of a hollow-fiber loose nanofiltration (HF-LNF) membrane process for potable water production was evaluated for an aerated, organic-laden, hard Florida groundwater containing slime bacteria. The HF-LNF process experienced no detectable fouling when sand filtration pretreatment was employed, and the membrane's durability and performance remained unchanged while producing a consistent water quality for 2,074 h of pilot run-time. Results of the study showed that a HF-LNF membrane with a 1,000 Dalton cut-off removed turbidity, sulfate, total organic carbon (TOC) and color by an average of 82%, 10%, 25%, and 95%, respectively. A decrease in permeate back pressure from 100 to 0 psi (6.9 to 0 bar) resulted in an increase in the normalized specific flux from 1.2 to 1.5 gfd/psi (29.6 lmh/bar). This decrease of nearly 87% in operating pressures did not significantly (<3%) affect the membrane removal efficiency for turbidity, sulfate, nor TOC. Construction and operating costs for a 2 million gal/d (permeate) HF-LNF process employing sand filtration for pretreatment and operating at an 85% water recovery rate was estimated to be $4.4 million and $0.57/kgal, respectively.

Keywords: Hollow fiber; Nanofiltration; Ultrafiltration; Capital costs; Operating costs

1. Introduction

Low-pressure hollow-fiber ultrafiltration (HF-UF) membrane processes are now widely used to remove turbidity and particles while providing a barrier for pathogens such as *Cryptosporidium* and *Giardia* during potable water production [1]. Unlike HF-UF, however, spiral-wound nanofiltration (SW-NF) membrane processes, developed in the 1980s, can achieve hardness, TOC, and synthetic organic compound removal [2–6]. As compared to a HF configuration, SW membranes are more energy intensive, prone to fouling, and often required advanced pretreatment for surface water supplies [6–8].

While the transition from bench-scale to full-scale plant implementation has occurred for SW-NF technologies, advancements in HF-NF had been historically limited to bench-scale applications. A number of HF-NF bench-scale studies were completed that documented the combined chemical resistance and organic retention of HF-NF via the study of membrane packing density, surface area to volume ratios, and self-support capabilities [9–12]. Comparisons between HF-NF and flat sheet membrane pretreatment requirements were also reported by Van der Bruggen and colleagues using similar bench-scale methods [13]. Additional bench-scale studies by Fang, Shi and Wang further demonstrated the applicability of HF-NF membranes for cost-effective natural organic matter (NOM) removal [14]. Sun and researchers have been successful in crosslinking polyethyleneimine (PEI) on polyamide-imide hollow fiber NF membranes to reduce pore size and increase the hydrophilicity of the membrane surface [15,16]. The PEI-modified membranes showed great potential in treating highly concentrated wastewater from the dye manufacturing industry in bench and pilot scale applications [17–19].

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More recently, pilot-studies that have been conducted evaluating the performance of HF-LNF as a prelude to full-scale surface water treatment application, primarily focused on NOM and color removal. Knops and colleagues showed HF-LNF technology was successful at removing dissolved constituents of a surface water supply in the Netherlands [20]. Linden and Persson compared HF-UF and HF-NF at the pilot-scale for the direct filtration of three alternative surface waters and found NOM removal on average was 55% and 88%, respectively, for each technology [21]. Kohler and colleagues performed a 6-month pilot evaluating HF-NF in short-term runs (<2 h) for direct filtration of surface water and found that an average of 88% of the NOM in the source water could be removed [8]. In a similar fashion, Keucken and colleagues piloted capillary modified polyethersulfone NF membranes treating soft, organic-laden (NOM = 7.0–10.0 mg/L) surface water (SUVA = 2.7–3.3 mg/L) at 80% water recovery in the northern part of Sweden [22]. Under these conditions, 70% of the NOM was removed (from 8 to below 2 mg/L). However, only 40% of the low molecular weight acids (MW between 300 and 400) were retained. The authors performed an advanced autopsy of the HF-LNF membranes after 12 months of operation and found no substantial changes that could be identified with the membrane morphology.

Although there has been a number of bench and pilot HF-LNF studies presented in the literature, it appears that few cost evaluations have been performed regarding HF-LNF treatment, and those that have been identified have been limited to surface water applications. No known groundwater treatment applications have been published in the literature. Sethi and Wiesner did simulate the treatment and cost effectiveness of HF-NF as compared to an integrated system comprised of HF-UF and SW-NF treating surface water using numerical simulation and cost modeling [23]. HF-NF membranes were simulated to handle higher concentrations of particulate and colloidal fouling as compared to SW-NF membranes which was assumed to require HF-UF as pretreatment. The cost of direct HF-NF treatment as compared to the integrated HF-UF and SW-NF processes was found to offer significant cost savings (approximately 30%) for small scale treatment plants (< 1 million gallons per day [MGD]; or <3,785 m^3/d). Additionally, Linen and Persson, using pilot-plant data, demonstrated that HF-NF process costs were approximately 30% more than the total cost of a conventional drinking water plant treating surface water [24]. It does not appear that costs have been developed for the HF-LNF technology for ground water treatment applications.

Although several HF-LNF membrane facilities have been constructed and are operating in northern Europe, the facilities tend to be small-scale (<1 MGD; or 3,785 m^3/d) and are designed to treat for NOM and color [20]. Ongoing research into HF-LNF continues but is focused with regards to surface water supplies and not generally groundwater supplies. This article presents a cost and performance study of a pilot-scale HF-LNF application for treatment of a biologically-active aerated groundwater supply containing high levels of sulfate, hardness, and dissolved NOM.

2. Experimental

2.1. Pilot location

The HF-LNF pilot was located at the City of Sarasota's (City) water treatment facility (WTF), located at 1750 12th Street in Sarasota, Florida, where water originating at the Verna wellfield is disinfected. The City relies on a natural-draft tray-aeration system for carbon dioxide and hydrogen sulfide removal from the groundwater at the Verna wellfield; after aeration the Verna water is chlorinated for biological control and piped approximately 20 miles (32 km) to the City's WTF. The aerated water is either treated using IX or bypassed and blended prior to disinfection with chorine. Approximately 5.2 MGD (19,684 m^3/d) of Verna groundwater is treated by cation exchange. The Verna groundwater contains elevated levels of hydrogen sulfide, hardness, and sulfate as seen from historical data provided in Table 1 (Tharamapalan, 2014) [25]. The Verna water source contains approximately 450 mg/L as $CaCO_3$ of hardness originating from the dissolution of underground rock formations such as calcite ($CaCO_3$), and dolomite ($CaMg(CO_3)_2$) [26].

2.2. Procedures

2.2.1. HF-LNF membrane pilot system

The HF-LNF membrane pilot equipment was provided by Pentair X-Flow (Marssteden 50, NL-7547 TC Enschede, Netherlands) and incorporated one pre-assembled filtration unit, consisting of a feed pump, a strainer, one membrane module, and an electrical cabinet. Associated piping and additional appurtenances such as pressure gauges, flow valves, and flow meters were included in the system. The system power requirements were 230 V, 50 Hz, and 25 A and consumed an average of 2.95 kW during the study. The process requirements for the system were designed to operate under the following conditions: operating pressures between 0 and 125 psi (0 and 8.6 bar); operating ambient temperature between 0 and 40°C, acidity of the medium between 3 and 11 pH units, and a flow rate of 2.64 gpm (0.6 m^3/h). The membrane module was equipped with Pen-

Table 1
Historical verna water quality

Parameter	Historical average
Alkalinity, mg/L as $CaCO_3$	159
Bromide, mg/L	0.06
Calcium, mg/L	91
Chloride, mg/L	18
Conductivity, µS/cm	935
Color (true), CPU	3.0
Magnesium, mg/L	135
Sulfate, mg/L	405
Sulfide, mg/L	6.2
TDS, mg/L	846
TOC, mg/L	2.00
UV-254, cm^{-1}	0.03

tair's HFW-1000 membrane, referring to an inside-out hollow fiber membrane with a 1000 Dalton molecular weight cut-off. The membrane is hydrophilic and composed of PES/modified PES. The HFW-1000 lumens have hydraulic diameters of 0.8 mm contributing to a total membrane area of 430 ft^2 (40 m^2).

A portion of testing the HF-LNF membrane pilot required the use of a sand filter for pretreatment. The sand filtration system was constructed of a filament-wound fiberglass filter with a 56 in (142 cm) sidewall length and housed (12 in) 30.5 cm of 1/8 in to 1/4 in gravel and (33 in) 83.8 cm of 0.45 to 0.55 mm fine silica sand. The sand filter was operated in a declining rate down-flow filtration setting using the pressure head (approximately 20 psi; 1.38 bar) available in the Verna pipeline. Originally the HF-LNF pilot system was designed to operate at a constant flux of 9 gfd (15 lmh) and a 50% recovery by providing back pressure on the permeate stream to maintain the recovery, allowing for adjustments in the filtrate, concentrate, and feed flows. The HF-LNF membrane pilot was evaluated at multiple settings to determine feed pressure requirements, operating flux values, and pretreatment requirements. Each setting was operated for a minimum of two weeks. SF was implemented as a pretreatment process for the initial operation of the HF-LNF pilot system, and was eventually bypassed for the last 600 h of operation to determine membrane performance without pretreatment. Water quality monitoring included the collection of pH, temperature, conductivity, total suspended solids (TSS), total dissolved solids (TDS), turbidity, alkalinity, sulfate, color and TOC. Hydraulic parameters that included flows and pressure were manually recorded three times a day. Samples representative of the aerated raw Verna (RV) water, SF filtrate, HF feed, HF concentrate, and HF permeate were taken at the corresponding sampling points shown in Fig. 1.

3. Results and discussion

3.1. Pilot-scale HF-LNF membrane testing using aerated groundwater

The single module pilot was operated in a cross-flow configuration under seven different settings. Table 2 provides a summary of the feed pressure, recoveries, and pretreatment requirements assessed during the pilot testing phase of this research.

3.1.1. Time utilization and operating conditions

The operation of the HF-LNF pilot commenced on June 25th, 2013 and was operated with minimal interruptions until final shutdown occurred on October 1st, 2013. The total runtime included 2,074 h over the course of nearly 100 d. Runtime refers to the time the pilot is either in forward filtration mode, backwash mode, or chemical enhanced backwash (CEB) mode. Although backwashes were not required during the duration of this study. Time utilization does not include the time it took to perform pressure decay tests, clean in place events, or additional pilot maintenance. Pilot downtime was experienced due to one scheduled event and two unavoidable events. The distribution of runtime and downtime is provided in Fig. 2. The first downtime event was strategically planned and occurred between July 5th and July 11th, 2013. The pilot design supplied by the manufacturer was constrained hydraulically and modifications were conducted to incorporate a more versatile design, which allowed for recoveries, water production, and pressures to be varied.

On July 19th, 2013 and a runtime hour of 428, the HF-LNF pilot data logger and acquisition system ceased to operate resulting in four days of hydraulic data loss. The HF-LNF pilot was taken offline to troubleshoot the data logger and repair piping which had experienced localized leaks during operation. The HF-LNF pilot resumed sta-

Table 2
HF-LNF pilot settings

Setting	Recovery	Pressure (psi; *bar*)	Pretreatment
1	50%	High (120;*8*)	Sand filter - Strainer
2	50%	Moderate (60;*4*)	Sand filter - Strainer
3	50%	Low (15;*1*)	Sand filter - Strainer
4	75%	Low (15;*1*)	Sand filter - Strainer
5	85%	Low (15;*1*)	Sand filter - Strainer
6	50%	Low (15;*1*)	Strainer
7	85%	Low (15;*1*)	Strainer

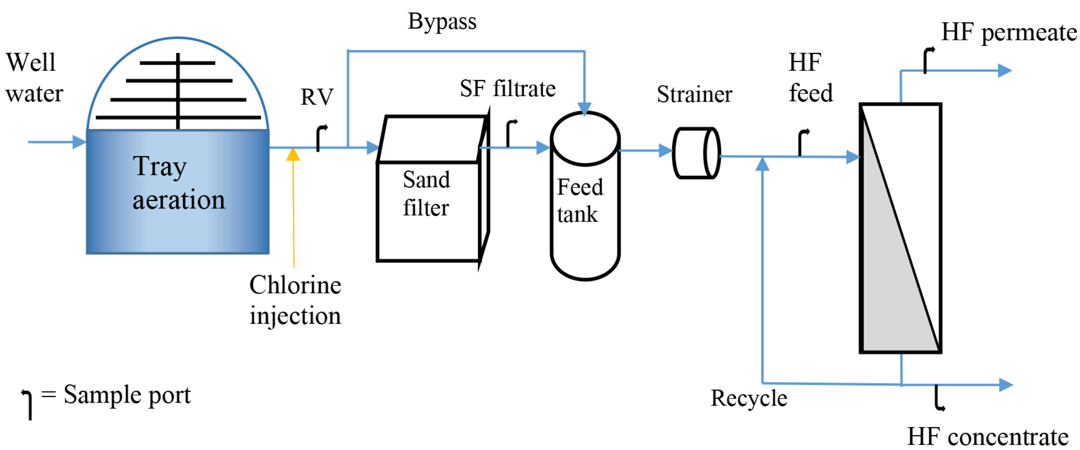

Fig. 1. Process flow diagram of HF-LNF membrane pilot.

ble operation on July 23rd, 2013 after piping repairs were complete. Unfortunately, repairs to the data logger were unsuccessful therefore manual readings were conducted 3–4 times daily for the remainder of the pilot testing experiments. The maintenance hours on the data logger and piping along with the accumulation of sand filter backwash hours accounted for a total of 130 h of unplanned downtime.

3.1.2. System water production and water recovery

The SF pilot operated in declining rate rapid filtration mode and was backwashed when necessary to produce sufficient feed water to the HF-LNF pilot. Backwashes were typically conducted on a weekly basis for a duration of 30 min. The sand filter was operated in down-flow filtration during normal operations and up-flow filtration during backwashes; additionally, the sand filter filtrate was used as the feed to the HF-LNF pilot and the backwash water was directed to the City's wastewater system. The HF-LNF pilot was operated in an inside-out cross-flow configuration with up-flow filtration and an internal recycle stream. The pilot had the ability to perform backwashes using down-flow filtration if cleanings where necessary.

Traditional SW-NF membranes operate at feed pressures of approximately 100 psi (6.9 bar). In contrast, typical feed pressures for HF-UF systems operate in the 4 psi (0.28 bar) to 22 psi (1.5 bar) range [27]. The pressures for the feed, concentrate, and permeate streams were recorded and plotted to produce Fig. 3. The first setting on the HF-LNF pilot operated at an average feed pressure of 120 psi (8.3 bar), a conservative permeate flow rate of 3 gpm (0.8 m^3/h), a recovery of 50%, and incorporated a back pressure of approximately 110 psi (7.6 bar) on the permeate stream.

Modifications to the pilot unit were performed which increased the functionality of the HF-LNF pilot by eliminating back pressure and consequently reducing operating pressures. Pressures were reduced in two intervals referred to as settings 2 and 3. Throughout the duration of settings 1 and 2 the flow measurements were validated by manually performing timed bucket tests. By the end of setting 2, modifications were performed on the pilot to allow the

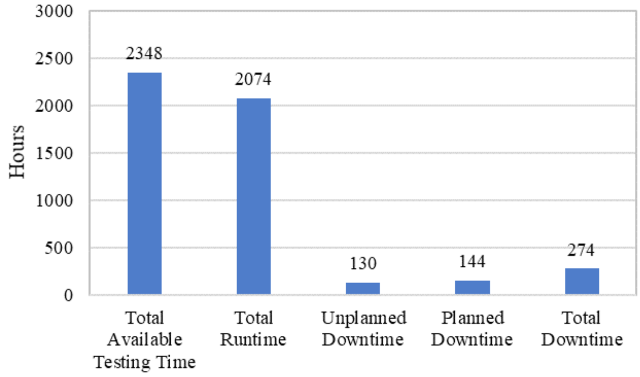

Fig. 2. Distribution of total available runtime and downtime events for HF-LNF pilot.

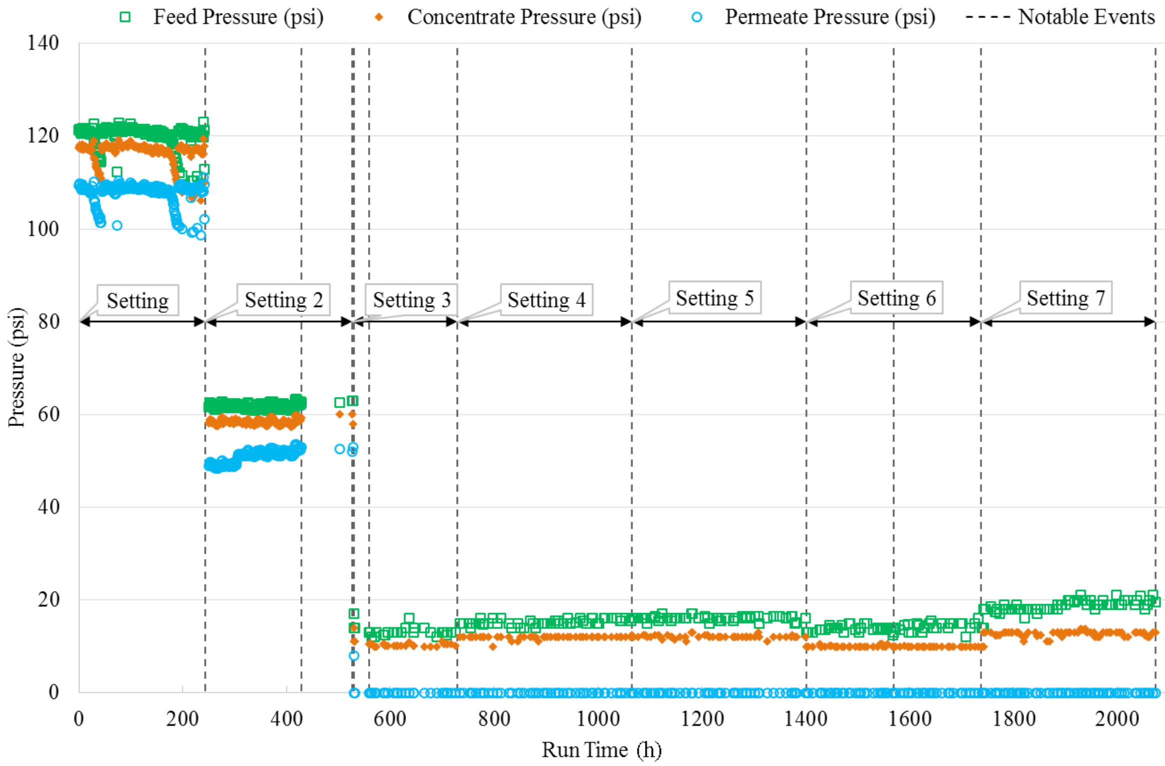

Fig. 3. HF-LNF pilot operating pressure requirements.

system parameters to be monitored visually by inspecting the analog flow meters and pressure gauges. Testing under setting 2 conditions began at runtime hour 243 as depicted by the vertical dashed line in Fig. 3.

Decreasing the back pressure resulted in a 50% reduction (60 psi; 4 bar) in feed pressure of the system. The pilot was tested under these conditions until a runtime of 527 h. The data logger was damaged during this setting as seen from the absence of pressure data in between the runtimes of 428 and 527 h of Fig. 3. After attempts to replace the data logger failed, operation resumed and manual recordings were initiated. At a runtime of 530 h, the third pressure adjustment (setting 3) was completed reducing the back pressure to atmospheric conditions and the feed pressure to 15 psi (1 bar).

The remaining experiments were conducted without implementing permeate back pressure, which decreased operating pressures and energy requirements. Settings 4 and 5 operated at permeate flow rates of 4 and 5 gpm (0.9 and 1.1 m^3/h) and recoveries of 77 and 85%, respectively. Feed pressure requirements for settings 4 and 5 were approximately 16 psi (1.1 bar). During setting 6 the recovery was decreased to 50% and the sand filter removed as a pretreatment step. Feed pressure requirements decreased and returned to 14 psi (0.96 bar) which was similar to pressure requirements required under setting 3 conditions. The final setting targeted a recovery of 85% with a permeate flow rate of approximately 5 gpm (1.1 m^3/h). Operating without sand filter pretreatment caused operating feed pressures to increase to approximately 19 psi (1.3 bar). The average operating pressures and flows for each setting have been provided in Table 3. Recoveries of 50%, 75%, and 85% were targeted in this research but differed slightly due to the mechanical operation and variability of the pilot controls.

3.2. Water flux and mass transfer coefficient

Flux values of 10, 13, and 17 gfd (17, 22, and 29 lmh) were achieved during operation of the pilot. The flux, feed temperature, transmembrane pressure (TMP), and normalized specific flux for the pilot study are shown in Fig. 4. During the first 383 h of runtime, the pilot was operated to produce a water flux of 10 gfd (17 lmh). The TMP during the operation of setting 1 was approximately 10 psi (0.7 bar). Setting 2 operated with an average TMP value of approximately 11 psi (0.8 bar) until a correction to the permeate back pressure was performed at a runtime of 306 h. The abrupt decline in TMP to 8 psi (0.5 bar) was not indicative of fiber breakage but rather a change in the operation. During setting 3, permeate production was increased to 4 gpm (0.9 m^3/h) yielding a flux of 13 gfd (22 lmh) and a TMP increase of 2 psi (0.1 bar). The reduction of back pressure on the permeate stream from 100 psi (6.7 psi) to atmospheric conditions resulted in an increase in the normalized specific flux as seen through settings 1–3. This indicates that the pilot system can be operated at lower pressures without having a significant effect on membrane permeate production. Settings 4 and 5 operated at a flux of 17 gfd (29 lmh) but the pilot experienced a slight increase in the TMP and decrease in specific flux corresponding with the system's recovery adjustment which occurred at runtime hour 1,065.

Operating conditions for setting 6 were similar to setting 2, that is, a recovery of 50% and a flux of 13 gfd (22 lmh) but without SF pretreatment. During setting 6 at runtime hour 1,570, a decrease was observed in the flux that corresponded to a slightly lower value than anticipated for the originally planned set point. The pilot was adjusted to correct the difference in flux from 12 gfd to 13 gfd (20–22 lmh) as denoted in Fig. 4 by the vertical dashed line. Setting 7 was also operated without the use of SF pretreatment. TMP increased to an average of 16 psi (1.1 bar) while operating

Table 3
Averaged HF-LNF pilot parameters for each testing setting

Setting	Runtime (h)	Feed pressure (psi) *bar*	Concentrate pressure (psi) *bar*	Permeate pressure (psi) *bar*	Feed flow (gpm) *m^3/h*	Recycle flow (gpm) *m^3/h*	Concentrate flow (gpm) *m^3/h*	Permeate flow (gpm) *m^3/h*
1	0–243	120	117	108	49.5[b]	44.2[b]	2.4[a]	2.9
		8.3	*8.1*	*7.4*	*11.2*	*10.0*	*0.5*	*0.6*
2	243–527	62	58	51	49.5[b]	44.1[b]	2.4[a]	3.0
		4.3	*4.0*	*3.5*	*11.2*	*10.0*	*0.5*	*0.7*
3	527–729	13	10	0	49.5	41.7	3.9	3.9
		0.9	*0.7*		*11.2*	*9.5*	*0.9*	*0.9*
4	729–1,065	15	12	0	50.5	44.1	1.5	4.9
		1.0	*0.8*		*11.5*	*10.0*	*0.3*	*1.1*
5	1,065–1,401	16	12	0	50.7	44.8	0.9	5.0
		1.1	*0.8*		*11.5*	*10.2*	*0.2*	*1.1*
6	1,401–1,738	14	10	0	48.7	40.9	4.0	3.8
		1.0	*0.7*		*11.1*	*9.3*	*0.9*	*0.9*
7	1,738–2,074	19	13	0	47.9	42.2	0.8	4.9
		1.3	*0.9*		*13.1*	*11.5*	*0.2*	*1.3*

[a]Concentrate flows were determined using bucket tests
[b]Feed flow assumed to be constant due to pilot limitations

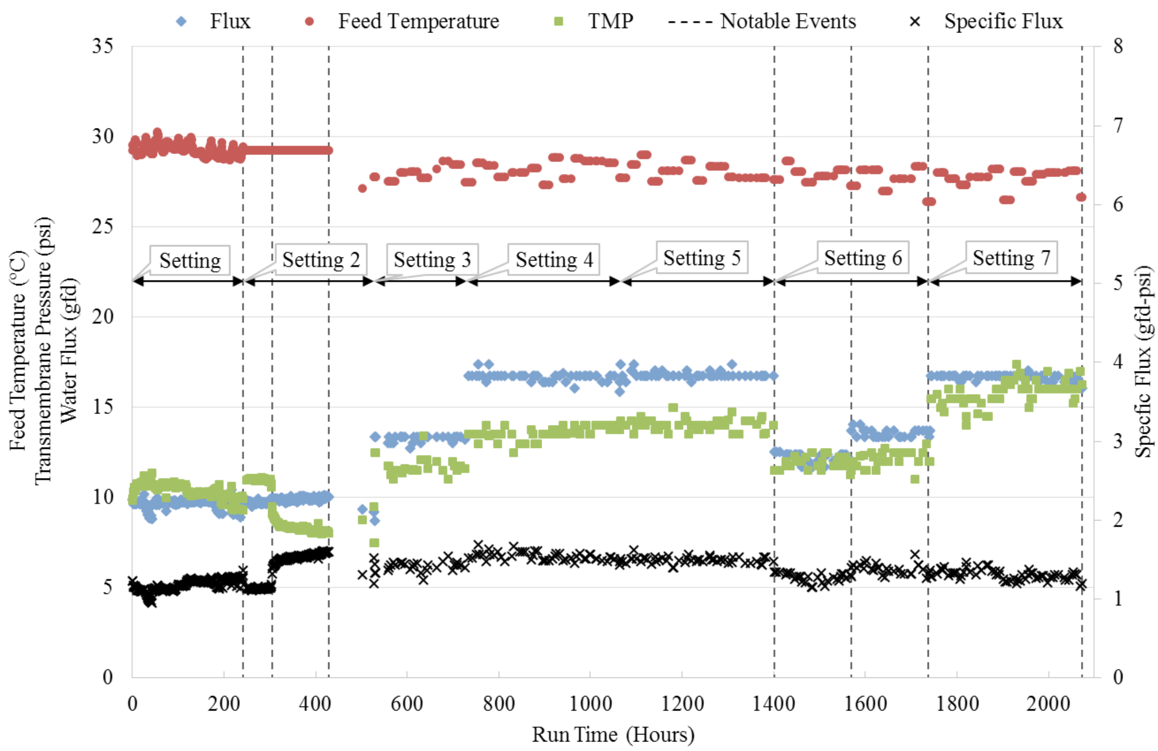

Fig. 4. HF-LNF pilot operating conditions.

at a flux of 17 gfd (29 lmh) without pretreatment. TMP varied over the course of the study from approximately 8 psi (0.5 bar) during setting 2 to a maximum of 17 psi (1.2 bar) during setting 7.

The average hydraulic parameters for each setting including recovery, water flux, TMP and temperature corrected water mass transfer coefficients (MTCs) have been listed in Table 4. The operating conditions for settings 6 and 7 were compared to settings 2 and 5, to assess the effect the SF pretreatment had on the hydraulic parameters of the membrane. Average TMP and specific flux values between settings 2 and 6 were similar indicating that fouling did not occur while operating at a flux of 13 gfd (22 lmh). On the other hand, hydraulic comparisons between settings 5 and 7 showed a 14% increase in TMP when operating at a flux of 17 gfd (29 lmh), but significant trends in specific flux or TMP were not apparent in the data therefore implementation of backwashes and cleanings were not necessary to maintain stable operation.

The MTC of water for a single stage membrane system can also be estimated using the homogenous solution diffusion (HSD) theory by plotting the water flux versus the transmembrane pressure differential as shown in Fig. 5 [28]. The MTC was determined from the slope of the x-y scatter plot using linear regression as 0.94 gal/sfd-psi or 0.054 d^{-1}. The coefficient of determination, R^2, was determined to be 0.989 indicating that nearly 99% of the variation could be described by the regression line. The number of observations in the linear model was 60 data points. The root mean square error (RMSE), sum of squares for error (SSE), and total sum of squares (SST) and were calculated to be 1.5, 127, and 11949, respectively.

Table 4
Calculated hydraulic parameters averaged for each testing setting

Setting	Runtime (h)	Recovery	TMP (psi) *bar*	Flux (gfd) *lmh*	Normalized specific flux (gfd/psi) *lmh/bar*
1	0–243	50%	10	10	1.2
			0.7	*17*	*29.6*
2	243–527	54%	9	10	1.4
			0.6	*17*	*29.6*
3	527–729	50%	12	13	1.4
			0.8	*22*	*29.6*
4	729–1,065	77%	14	17	1.5
			1.0	*29*	*29.6*
5	1,065–1,401	84%	14	17	1.5
			1.0	*29*	*29.6*
6	1,401–1,738	49%	12	13	1.3
			0.8	*22*	*29.6*
7	1,738–2,074	86%	16	17	1.3
			1.1	*29*	*29.6*

3.2.1. System water quality

The HF-LNF pilot system was sampled a minimum of 5 times per week for water quality analysis. Sample locations included the aerated Verna water, sand filter filtrate,

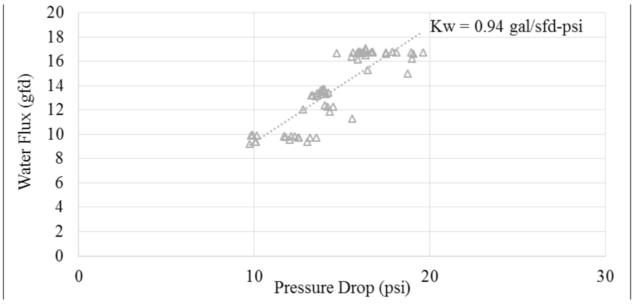

Fig. 5. Water flux vs. transmembrane pressure differential.

membrane feed, concentrate, and permeate. The HF-LNF membrane consistently produced permeate with turbidity values less than 0.2 NTU regardless of the SF pretreatment. On average, the membrane achieved 82% turbidity removal indicating rejection of suspended solids and particles in the feed stream. The pH of the water did not change significantly across the membrane nor was alkalinity significantly affected. The suspended solids concentrations in the Verna water varies throughout the pilot testing duration reaching a concentration as high as 6 mg/L; suspended solids were consistently removed from the permeate stream. The average TDS of the Verna water was determined to be approximately 800 mg/L, and varied from 720–950 mg/L.

The variation of TDS in the feed water was likely due to the operation of multiple wells with varying water quality from the Verna wellfield. The average maximum and minimum feed TOC concentrations ranged from approximately 2.2–2.5 mg/L. TOC removal was not affected by the removal of the pretreatment process, nor was it significantly affected by variations in flux or recovery. On average 25% of the TOC was removed using the membrane.

Anions analysis included measurements for sulfate. Although the average feed sulfate concentration was 430 mg/L the maximum and minimum feed sulfate concentrations varied from 322 mg/L to 500 mg/L, respectively. The HFW1000 membrane achieved a sulfate removal of 10% at 17 gfd (29 lmh). The water quality results are believed to indicate the removal mechanism of the HF-LNF membrane was predominately due to size exclusion, with the ability to remove TOC, color and turbidity but was not completely effective at removing sulfate.

4. Cost

Two alternatives were evaluated in this analysis: (i) a traditional SW-NF membrane process and (ii) a SF and HF-LNF process. The first treatment alternative, provided in Fig. 6, considered the use of traditional SW-NF membrane process with SF and HF-UF processes required for pretreatment.

The second treatment alternative investigated in this research considered HF-LNF membrane technology with SF pretreatment. Although the hydraulic performance of the HF-LNF pilot was not significantly affected with the removal of the SF pretreatment, indicating the membrane could possibly be used to treat Verna water without additional pretreatment, it was included in the cost estimates to

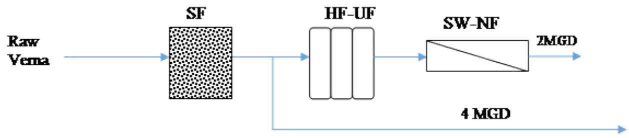

Fig. 6. Verna water supply treatment alternative 1 –SF, HF-UF, SW-NF.

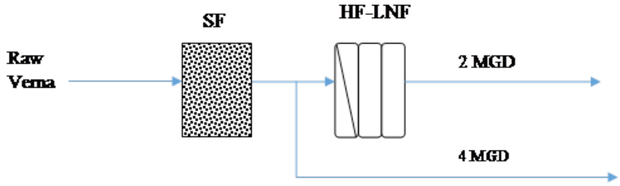

Fig. 7. Verna water supply treatment alternative 2 – SF, HF-LNF.

prolong membrane life. The second treatment alternative is shown in Fig. 7 and includes the filter and permeate flows. The conceptual costs presented herein were estimating assuming recoveries of 97%, 95%, and 85% for media filtration, membrane filtration, and membrane softening processes, respectively.

4.1. Capital costs

The capital costs for the Verna water treatment alternatives is provided in Table 5 in U.S. dollars 2018. The costs for additional buildings, degasifiers, clearwells, transfer pumps, ground storage tanks, bulk chemical storage, emergency power, yard piping, and site development have been excluded from the conceptual cost estimations. The direct capital costs for a typical rapid SF process would include horizontal pressure filters, filter under drains and distributors, air wash configurations, tank nozzles and manways, face piping, instrument and controls (I&C), and filtration media. Direct capital costs for SF equipment were developed using data from an existing facility (Jupiter, 2007). The capital costs of the SF process for the WTF in Jupiter FL were adjusted for plant size by a factor of 0.09. The conceptual capital costs for a 6 MGD SF process was estimated to be approximately $602,000.

The direct capital cost for HF-UF equipment includes HF-UF membrane modules, cleaning equipment, high pressure pumps, backwash pumps, transfer pumps, backwash tanks, blowers, I&C. The capital costs were calculated using membrane filtration cost curves found in the literature [29]. Capital costs for the HF-UF equipment were based upon a permeate flow of 2 MGD and were determined to be a total of $1.04 million.

The direct capital costs for SW-NF equipment includes SW-NF membrane modules, cleaning equipment, high pressure pumps, pretreatment chemical feed and storage, cartridge filters and I&C. Capital costs for estimating membrane softening processes were developed using cost curves similar to the HF-UF cost curves which consider the process capacity (MGD) as a function of cost ($/gpd) [29]. The direct capital costs for a SW-NF process were estimated to be approximately 2.4 million dollars.

Table 5
Conceptual capital costs for city's Verna treatment alternatives

Category	Alternative 1 Cost ($1000)	Alternative 2 Cost ($1000)
Direct Capital Costs		
Media Filtration Equipment Cost (horizontal pressure filters, filter underdrains and distributors, air wash configuration, tank nozzels and manways, face piping, instrumentation and controls, 16" gravel media, 24" sand media)	602	602
Membrane Filtration Equipment Cost (HF-UF membrane modules, cleaning equipment, feed/permeate pumps, backwash pumps, blowers, backwash tanks, chemical feed and storage, instrumentation and controls)	1,040	n/a
Membrane Process Cost (membrane modules, vessels and supports, cleaning equipment, feed pumps, pretreatment chemical feed and storage, cartridge filters, backwash pumps and blowers [not applicable for SW-NF], instrumentation and controls)	2,400	2,600
Total Direct Capital Costs	4,042	3,202
Indirect Capital Costs		
Construction Overhead and Profit (22%)	889	704
Insurance and Bonding (3%)	121	96
Contingencies (15%)	606	480
Total Indirect Capital Costs	1,617	1,281
Total Estimated Capital Costs	5,658	4,482
$/gallon/day Capital Installed	2.49	1.96

Indirect capital costs considered construction overhead and profit, insurance and bonding, and contingencies for each treatment alternative and were estimated to be 22%, 3% and 15% of the total direct capital costs, respectively. The total capital cost for the treatment alternative utilizing SF, HF-UF, and SW-NF was estimated to be approximately $5.7 million.

The capital costs estimates for the second treatment alternative utilizing SF and HF-LNF were estimated using SW-NF and HF-UF equipment conceptual estimates. The equipment required for HF-LNF treatment would likely include components from both treatment alternatives. For instance the addition of blowers and backwash pumps would include additional costs to the HF-LNF process. Previous cost estimates conducted by Sethi and Wiesner found HF-NF to be comparable to a UF-SW-NF system depending on plant size, economies of scale, and operating conditions. The conceptual capital cost for the HF-LNF equipment was estimated to be $2.6 million corresponding to a total capital cost of approximately $4.5 million but the capital costs of the HF-NF membranes are expected to decrease as the fabrication process improves and the technology advances [23]. In fact, Xia and researchers recently made a cost-effective dual-layer membrane by using multiple polymers during the fabrication process [32,33].

The total capital costs for each treatment alternative are provided in Table 6. The estimated installed conceptual capital cost for the treatment alternative using SF and HF-LNF was determined to be $1.96/gpd. The treatment alternative using SW-NF required an additional pretreatment process resulting in an increase of approximately $1.2 million, or $0.53/gpd, for a total estimated conceptual cost of $2.49/gpd.

Table 6
Conceptual capital process costs for each treatment alternative

Process (size)	Alternative 1 cost($/gpd)	Alternative 2 cost($/gpd)
Media filtration (6 MGD)	0.14	0.14
Membrane filtration (2 MGD)	0.67	–
Membrane softening (2 MGD)	1.68	1.82
Total cost ($/gal/d capital installed)	2.49	1.96

4.1. Operating and maintenance costs

The operating and maintenance (O&M) costs for each membrane treatment alternative would include: energy and power, chemicals, cartridge filter replacement, membrane replacement, water and sewer charges, cleaning chemicals, maintenance and labor [30]. A significant portion of the energy costs for NF processes are from the operation of high pressure feed pumps [1]. Operating feed pressures for each alternative were monitored and averaged for each of the pilots to estimate power requirements of the feed pumps. The amount of energy to drive the feed pumps for each pilot were estimated using Eq. (1) which considers the pump pressure (P), pump efficiency (η_p), motor efficiency (η_m), and recovery (R).

$$\frac{kwh}{kgal\ permeate} = \frac{P(psi) \times 0.00728}{\eta_p \times \eta_m \times R} \quad (1)$$

Each of the pilots feed pumps was assumed to have a pumping efficiency of 60% and a motor efficiency of 94%. The average industrial electricity rate in the City was determined to be approximately $0.07/kWh [31]. Additional costs from operating labor wages and fringes were not included assuming current plant personal could be utilized to operate the new treatment system. Furthermore chemical costs, administration and overhead costs were considered to be comparable for each treatment alternative. Costs for membrane replacement were calculated using the method proposed by Byrne assuming an average membrane life of five years [23,30]. Operating costs for the SW-NF process included cartridge filter replacement which was not included in the HF-LNF estimate. Both treatment alternatives neglect the cost of concentrate disposal, but it is noted that the City has two options for disposal including deep well injection or sewer. The conceptual O&M conceptual cost estimates for the full-scale alternatives are provided in Table 7.

The conceptual annual operating costs for the SW-NF treatment alternative were estimated to be $278,000 or approximately $0.11/gpd. The conceptual annual operating costs the HF-LNF alternative was estimated to be $0.06/gpd resulting in a yearly savings of approximately $107,000. Capital costs components listed previously were amortized over the design life of the membrane plant assumed to be 20 years with a 4.5% interest rate to determine the total (capital and O&M) annual costs provided in Table 8. The total cost for the SW-NF treatment alternative was estimated to be $278,000/year or $0.84/kgal. Alternatively the total amount for a new HF-LNF treatment alternative was estimated to cost approximately $170,000/year, or 0.57/kgal.

Table 7
Conceptual O&M costs for city's verna treatment alternatives

Category	Alternative 1 cost ($1000)	Alternative 2 cost ($1000)
O&M costs		
Energy and power	135	91
Chemicals	18	13
Membrane replacement	85	41
Cartridge filter replacement	5	–
Administration and supplies	6	4
Overhead (15%)	20	12
Miscellaneous	10	10
Total estimated annual operating cost	278	170
$/gallon/d O&M	0.11	0.06

Table 8
Total process cost summary for each treatment alternative

Process (size)	Alternative 1 cost ($/kgal)	Alternative 2 cost ($/kgal)
Media filtration (6 MGD)	0.07	0.07
Membrane filtration (2 MGD)	0.27	–
Membrane softening (2 MGD)	0.51	0.51
Total cost	0.84	0.57

5. Conclusions

Although SW-NF membranes have historically been used to treat organic-laden hard groundwater supplies, the process is energy intensive and requires HF-UF membranes to prevent fouling. HF membranes offering nanofiltration properties aim to integrate the hydraulic operation of a conventional HF-UF process while targeting the rejection capabilities of SW-NF membranes by combining the technologies. The results of this study provide performance data including hydraulic operations, fouling potential, and water quality rejection data for a pilot-scale HF-LNF membrane.

5.1. Hydraulic operations

By decreasing permeate back pressure from 100 to 0 psi (6.9 to 0 bar) the normalized specific flux increased from 1.2 gfd/psi to 1.5 gfd/psi (29.6 lmh/bar) which improved the productivity of the system. Furthermore a decrease of nearly 87% in operating pressures did not significantly (<3% difference) affect the membrane removal efficiency for the targeted constituents but would provide significant energy savings for full-scale systems.

5.2. Fouling potential

TMP, water flux, specific flux and water quality results indicate that the membrane did not experience significant fouling with or without sand filtration pretreatment. Additionally a pressure decay test (PDT) was conducted on the membrane at the conclusion of the study indicating no fiber breakage and no loss of performance.

5.3. Water quality

Pilot testing results show turbidity is effectively removed with permeate turbidity consistently less than 0.2 NTU. The membrane was successful at partially removing sulfate and TOC with 10 and 25% rejection, respectively. 95% of the TOC in the groundwater supply was comprised of dissolved constituents indicating partial removal of dissolved organic carbon.

The results of this study indicate that HF-LNF membranes could be operated under low pressure conditions offering hydraulic advantages and significant cost savings compared to SW-NF membranes. Conceptual capital costs for the HF-LNF treatment alternative and the SW-NF treatment alternative were estimated to be $4.4 million and $5.6 million, respectively. Excluding the cost of labor and fringes, conceptual operating cost for the HF-LNF process including SF pretreatment was estimated to be $0.57/kgal. The conceptual operating cost for the traditional SW-NF process including SF, HF-UF, and CF pretreatment was estimated to be $0.84/kgal.

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