Ceramic Ultrafiltration Membranes with Improved Economics, Operability, and Process Design Flexibility

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ABSTRACT

With the latest advances in ceramic ultrafiltration (CUF) membrane manufacturing, the capital costs are now competitive to polymeric hollow fiber UF (PUF) membranes. CUF membranes have longer life, are more robust, and have shown a freedom from operational limitations that plague PUF membranes, i.e., maintaining a high flux rate in cold water and sustainable operation in a wide range of water quality conditions. Case studies presented to highlight these features in a variety of applications.

INTRODUCTION

The production of safe drinking water requires suitable water supplies and purification processes to adequately provide safe drinking water in compliance with drinking water standards. The multi-barrier purification process approach frequently adopted by the USA and Canada often uses microfiltration or ultrafiltration membranes (MF/UF) as a physical barrier to protect the public from unsafe levels of microbiological contaminants. Industrial companies also require clean water for cooling, boiler makeup, and a variety of manufacturing processes. It is common to find these same MF/UF membranes applied to reduce turbidity and suspended solids of industrial process water directly, or in use as pretreatment to protect the performance of Reverse Osmosis (RO) membranes. As ground water and clean surface water sources become more scarce and restricted from use, the municipal and industrial users increasingly turn to impaired water sources or reusing treated waste water effluent to reduce their water shortage risks.

The use of traditional polymeric MF/UF membranes (e.g. PVDF, PES…) is established in the industrialized nations. MF/UF membrane systems have seen increasing use in recent years due to decreasing prices, resulting from manufacturing advances and technological improvements to packing densities and hollow fiber designs. Today, typical lifetime guarantees for polymeric MF/UF membranes range from 3-5 years for industrial water treatment systems and 7-10 years for municipal systems, which are more conservatively designed. Although improvements have been made in terms of fiber breakage and permeability losses over time, these issues still remain. In addition, polymeric membranes are limited in temperature resistance, resistance to solvents and other chemicals, suspended solids levels in feed stream, and have a low resistance to abrasion from sand or activated carbon. As the global demand grows to treat more challenging waters in reuse and waste water applications, the short comings of polymeric membranes will become even more acute.

Ceramic MF/UF membranes continue to make improvements in increasing surface area and lowering prices, resulting from innovative production techniques and a focus on water treatment applications. The invention of the high surface area monolith structure (Goldsmith, 1988) was a major breakthrough in ceramic module construction to reduce the cost and broaden the applications of ceramics. Additional new developments in ceramic membranes by Nanostone Water, Inc. (the Company) include a patented design (Göbbert and Volz, 2010) where individual ceramic segments are potted together forming the monolith structure, significantly reducing the production costs compared to other ceramic membrane modules. These advancements now show ceramic membrane systems to be cost competitive with polymeric MF/UF membrane systems.

Figure 1: High Surface Area Ceramic Membrane Module

In many cases, a ceramic MF/UF membrane system can operate with less pretreatment, such as eliminating the need for a clarification system which has a significant impact on the initial capital cost, operating cost, and foot print. With the higher total suspended solids (TSS) tolerance of ceramic membranes and the ability to utilize more aggressive chemical and hydraulic cleaning methods, the risk of irreversible fouling is less compared with polymeric MF/UF membranes. The high TSS tolerance of ceramic membranes also allows much greater flexibility in the process design of the system. For example, a lower capital cost solution can be achieved by designing the system with a high flux rate such as 200-300 Gallons per day per ft^2 of membrane (GFD) using higher coagulant dosage and more frequent maintenance cleaning intervals. This results in a higher operating cost in terms of energy and chemical consumption but with a lower up front capital cost. Conversely, a more conservative flux rate such as 100 GFD can be used with lower amounts of coagulant dosage and less frequent chemical cleaning to reduce the overall operating costs. The higher TSS tolerance also allows the system recovery rate to be increased to very high levels if needed for the project. With the higher tolerance levels of the ceramic membrane, the system design can be tuned to fit the needs of the project with a much wider range than polymeric UF membrane systems.

In cases with a cold water source < 40°F (< 5°C), the typical solution is to design the polymeric membrane facility at lower flux during colder weather periods which adds additional capital costs into the facility. The Company conducted testing of their ceramic UF membrane throughout all four seasons at a river water treatment plant where water temperatures range from 40°F to 80°F and the results are favorable. Ceramic membrane pores do not expand or contract inside these temperature ranges due to the materials of construction, and all other operational parameters are unchanged. The result is a competitive design flux that can be maintained throughout the year with only an increase in operating pressures that will follow a linear increase with water viscosity only.

PRETREAT LESS, RECOVER MORE

There are three main reasons contributing to a ceramic UF membrane's ability to operate with less pretreatment and/or at elevated recovery rates compared to conventional polymeric UF membranes: 1) Higher TSS tolerance; 2) Higher pressure limits; 3) Higher chemical resistance. With the higher TSS tolerance of the ceramic membranes, each filtration cycle can be extended, affording higher solids levels to accumulate on the membrane surface. In cases with high TSS in the feed water, an acceptable recovery rate can be achieved in most cases without requiring a clarification step. Secondly, the higher pressure limits afforded by ceramics allow for more aggressive hydraulic cleaning methods with higher pressures and flow rates over shorter durations. For example, the Nanostone CM-151™ module (ceramic UF membrane) is rated up to 100 psi (7 bar) of transmembrane pressure (TMP). Lastly, the inherent higher chemical resistance of ceramics enables a wide range of chemical cleaning routines to ensure stable permeability. With a much wider operating envelope and more aggressive cleaning options, recovery rates of up to 99% have been observed in some cases.

The higher ceramic recovery was validated in a comparative evaluation of the ceramic UF membrane and an existing polymeric UF system installed at an industrial waste water reuse facility. In an effort to accommodate fluctuations in water quality resulting in periods of high suspended solids loading, the polymeric UF system is operated at an average recovery rate of 90%. Applying the same feed water to a pilot system with the ceramic UF membrane, a sustained recovery rate of 97% was observed. Table 1, below, reports the annual water cost savings realized with the ceramic UF considering the facility's cost of source water and the 7% improvement in ceramic UF recovery compared to the polymeric UF.

Table 1: Cost of Water vs. Recovery Rate

	Polymeric UF System	Ceramic UF System
Flow	2.3 MGD*	
UF Recovery % RO Recovery % Overall Recovery %	**90%** 75% 68%	**97%** 75% 73%
Source Water Cost	$1.58 / 1,000 Gal	
Annualized Water Costs	$431,130	$361,487
Annual Water Consumption	273M Gal/ Yr.	229M Gal. / Yr.
Savings	**X**	**44M Gal. / Yr.** **$69,643 / Yr.**

*MGD = Million Gallons per Day

ADJUSTING DESIGN FLUX TO BALANCE CAPITAL AND LIFE CYCLE COST

In addition to allowing higher recovery and/or higher TSS in the feed water, ceramic membranes also allow very high sustainable design fluxes, compared to polymeric UF membranes. With moderate feed water TSS levels, the stable flux rates of ceramic membranes can be pushed to levels typically 3 to 5 times higher than what can be achieved with polymeric membranes. This allows the equipment design more freedom to balance upfront capital costs with longer term operating costs. In an effort to study the performance and economics of a system design with a wide range of operating fluxes, the Company conducted a long term pilot study on a direct river water feed source. During the study, the turbidity of the water was typically 3 to 10 NTU and the temperature ranged from 45 to 60°F (7-15°C). A 500 micron screen filter was used for basic pretreatment and a coagulant was added in line before the membranes with a 1 minute contact time. The coagulant selected was Aluminum Chlorohydrate (ACH) primarily because this was the same coagulant used on a nearby polymeric membrane drinking water plant operating on the same river water source and thus provides a good benchmark for performance.

The design of the polymeric membrane plant operating on this same water source was designed for a gross flux rate of 68 L/hr. per m^2 of membrane (LMH) or 40 GFD. This plant uses a self-cleaning strainer as the only pretreatment step and injects inline coagulation of 1 mg/L as Al^{+3} of ACH for flux enhancement and additional Total Organic Carbon (TOC) reduction.

The first flux level tested for the ceramic membrane was 115 GFD (196 LMH), also with a 500 micron self-cleaning strainer as pretreatment and 1 mg/L of ACH solution for flux enhancement, and additional TOC reduction just as with the bench mark polymeric membrane system. This represents a flux rate nearly 3 times the benchmark polymeric membrane system operating on the same water source with similar operating conditions. The ceramic system ran in dead end flow configuration with a backwash cycle every 20 minutes. The backwash flow rate was ~ 2 times the filtration flow or 230 GFD (400LMH) for a 10 second duration followed by a 15 second feed water flush at the same flow rate as normal filtration. This resulted in a recovery rate of >96%. A maintenance chemical clean in place (mCIP) was performed every 2 to 3 days. The mCIP process is the recirculation of cleaning chemicals through the feed channels of the membrane for 30 minutes or less. The typical mCIP chemicals used are sodium hypochlorite (NaOCl), sodium hydroxide (NaOH), and hydrochloric acid (HCl). Typically the NaOCl with additional NaOH added is the first step for mCIP; followed by a second step using HCl.

The results of the first flux level test are shown below in Figure 2 which plots the net driving pressure, also known as transmembrane pressure (TMP), over time. The TMP is in PSI and Bar of pressure and is normalized to 20°C or 68°F.

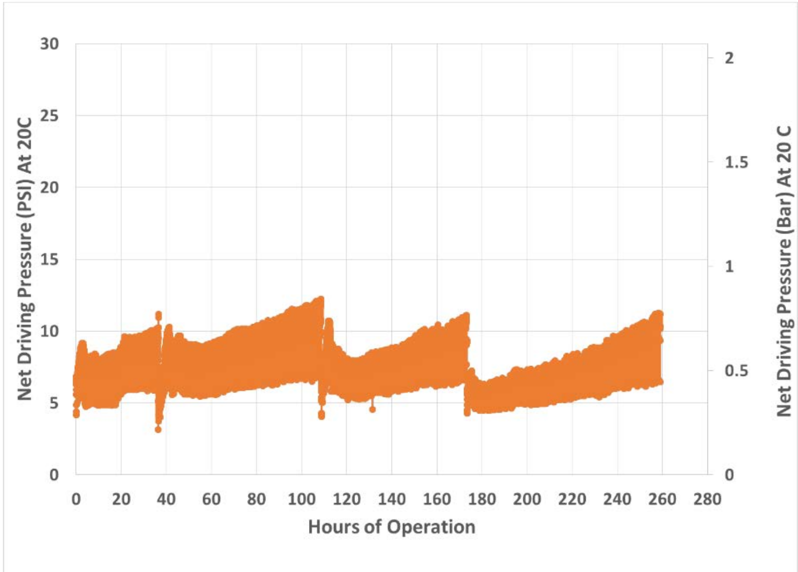

Figure 2: Pilot Test Results 115 GFD Flux Rate and 1 mg/L of ACH

The profile of the TMP is relatively flat at about 8 PSI (0.6 Bar), suggesting regular mCIP every 2-3 days will maintain the fouling control such that a full recovery chemical cleaning could be done on a scheduled basis – perhaps every 6 months to a year. The water quality during this test was between 3 and 10 NTU turbidity in the feed water and always < 0.08 NTU in the permeate, and typically at 0.05 NTU.

A series of flux increases above 115 GFD (196 LMH) were made in step fashion to evaluate the performance and to find a point of "critical flux" where the system becomes unstable. It was discovered that if the coagulant dosage was increased, there was a significant increase in stable flux levels that could be achieved. By raising the coagulant dosage to just 2 mg/L of ACH allowed the ceramic membrane to reach 184 GFD (313LMH). This is a flux level 4.6 times that of the benchmark polymeric membrane plant. In this test, the ceramic system ran again in dead end flow configuration, but this time with a backwash cycle every 15 minutes. The backwash flow rate was ~ 2 times the filtration flow or 368 GFD (626 LMH) for a 15 second duration followed by a 15 second feed water flush at the same flow rate as normal filtration. This resulted in a recovery rate of >95%. A maintenance chemical clean in place (mCIP) was performed every 2 to 3 days as before. The results of the second flux level test are shown below in Figure 3 which plots the net driving pressure or TMP normalized to 20°C or 68°F.

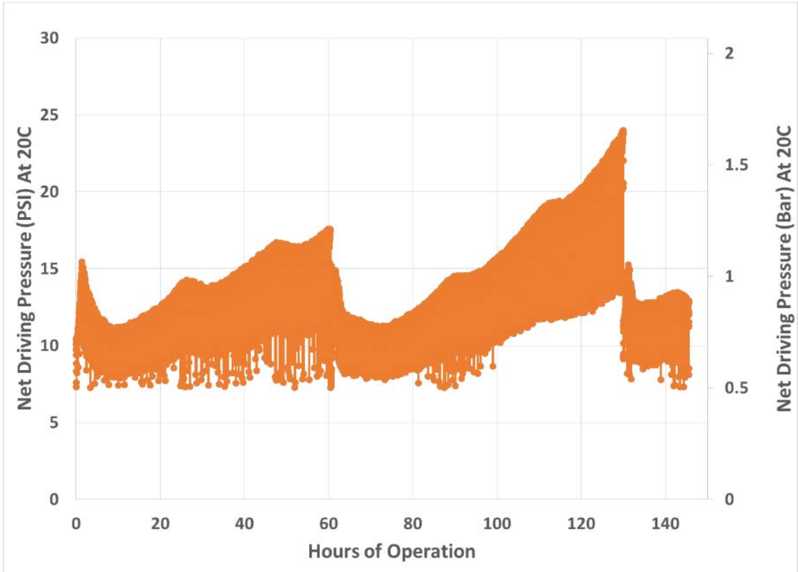

Figure 3: Pilot Test Results 184 GFD Flux Rate and 2 mg/L of ACH

The profile of the TMP here is stable and the mCIP step performed at day 3 and day 6 had a positive result in returning the TMP to the clean starting point. At this flux, the average TMP is about 14.5 PSI (1.0 Bar) and it is expected that regular mCIP steps will maintain a consistent pressure profile such that recovery CIP can be scheduled at 6 months to 1 year. The water quality during this test was about 3 NTU turbidity in the feed water and always < 0.08 NTU in the permeate, as before. Rather than continuing to run this flux level for several more days, it was decided to continue to make flux increases and experiment with a corresponding increase in coagulant to try to find the critical flux.

The experimentation of coagulant dosage and increasing flux ended at a flux rate of 230 GFD (391 LMH). The experiment ended ***not*** because the critical flux had been reached but because the capacity of the feed pump to the pilot system had been reached. This flux level at this point in the experiment was now 5.75 times higher than the 40 GFD benchmark polymeric UF system. The coagulant dosage at this flux level was 5 mg/L of ACH. In this test, the ceramic system ran again in dead end flow configuration with a backwash cycle every 15 minutes. The backwash flow rate was held at the 368 GFD (626 LMH) for a 15 second duration followed by a 20 second feed water flush at the same flow rate as normal filtration. This resulted in a recovery rate of >95%. A maintenance chemical clean in place (mCIP) was performed every 2 to 3 days, as before. The results of this third flux level test are shown below in Figure 4 which plots the net driving pressure or TMP normalized to 20°C or 68°F.

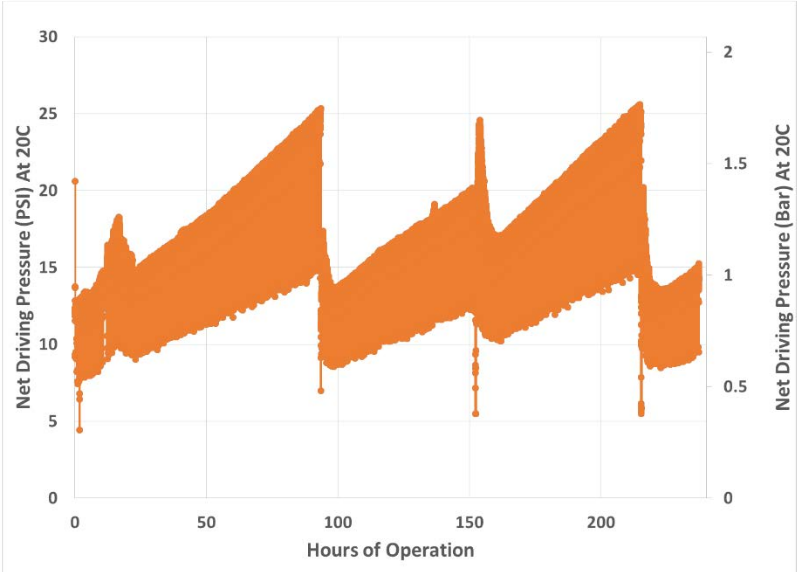

Figure 4: Pilot Test Results 230 GFD Flux Rate and 5 mg/L of ACH

The profile of the TMP here is also stable despite the very high flux rate. Here, again, the mCIP step performed at 3 day intervals has a positive result returning the TMP close to the start each time. At this flux the average TMP is about 17.4 PSI (1.2 Bar) and it is expected that regular mCIP steps will maintain a consistent pressure profile such that recovery CIP can be scheduled at 6 months to 1 year. The water quality during this test was the same as the previous test with a low 3 NTU in the feed water and < 0.08 NTU in the permeate.

An important fact with all of these tests, regardless of the flux level, is that the maintenance chemical cleaning is at an interval of 2 to 3 days. This means that if there were changes in the feed water quality simply increasing that mCIP interval to a 12-24 hour basis would allow the system to maintain a constant flux rate. With the additional chemical resistance of ceramic membranes, this increase in chemical exposure does not have the same risks of reducing permeability and life span of the membrane. Of course, the operator could simply reduce the flux and plant output, if they could afford to, during periods of upset conditions which is what is typically done with polymeric membranes as the primary action. The advantage of the ceramic membrane in these situations is that you have more choices as outlined in the table below.

Table 2: Operational Choices for MF/UF When Feed Water Quality Degrades

Polymeric MF/UF System	Ceramic UF System
1. Decrease output 2. Increase mCIP/CEB frequency; ***risking membrane life***	1. Increase coagulant dosage 2. Increase mCIP/CEB frequency ***without*** risking membrane life 3. Decrease output

With such a wide range of flux and coagulant dosage, the question arises on how the initial capital cost and longer term operating costs of ceramic UF membrane systems compare to each other as well as to a polymeric MF/UF membrane system as a benchmark. To consider these questions, a techno-economic model is constructed below to better analyze the capital and operating costs of the three different flux rates listed above, as compared to the benchmark polymeric MF/UF membrane plant. The main design and cost assumptions put into the model are shown in Table 3 below, as well as the capital cost estimates for each case. The final capital cost estimates are also shown graphically in Figure 5 below. For the example, a plant size of 2 MGD is selected. The basic assumption is that the capital cost of the primary components, not including membranes, is estimated at $0.50 / GPD. This is the same with the polymeric MF/UF plant as well as the ceramic membrane plant. This particular ceramic UF membrane process design uses conventional backwash pumps at moderate fluxes and so the same materials and basic design is the same as with pressurized polymeric MF/UF systems, and as a result, can fit into an “Open Platform” or “Universal Rack” concept where a number of membrane module suppliers can be used.

From this point there are credits or penalties given for more or less membrane modules charged at $2,000 USD per module. This compensates for the additional pipe manifolds, connections, couplings, labor, and frame material needed for a skid that has more membrane modules. The model then adds a penalty, as appropriate, for the cost of a higher pressure feed pump that would be needed for higher flux cases of the ceramic membrane. The model looks at the ratio of the clean TMP as compared to the polymeric UF design case. For the Ceramic UF case one scenario, the clean water permeability is about 3 times higher than the typical polymeric pressurized module. In this case, the flux of the ceramic UF is about 3 times higher and so the pressures will be about the same. For the ceramic UF case two, and case three, there is a penalty for the higher pressure pump required. There are a wide range of end user prices for polymeric MF/UF membranes depending on the size of the project, as well as whether the membranes are bid separately or included in a proprietary design from the manufacturer. A typical price of $35 USD/$M^2$ is used in this case as a typical value for end user prices in the initial purchase of the system.

Table 3: Techno-Economic Capital Cost Model Comparing a Typical Polymeric MF/UF Case to a Ceramic UF System at Three Flux Level Cases

Model Inputs	Polymeric MF/UF System	Ceramic UF System Case 1	Ceramic UF System Case 2	Ceramic UF System Case 3
Flow Rate (Max)	2.0 MGD, 7,571 m^3/day			
Water Source	Direct River Water, Design 15°C (59°F) Temperature, < 10 NTU Turbidity			
Flux at max flow Rate	40 GFD (68 LMH)	116 GFD (196 LMH)	182 GFD (309 LMH)	225 GFD (381 LMH)
Total Membrane Area Total Module #	50K Ft^2 (5K m^2) 64 Modules	17K Ft^2 (2K m^2) 66 Modules	11K Ft^2 (1K m^2) 42 Modules	9K Ft^2 (1K m^2) 34 Modules
Initial Capital Cost Excluding Membrane*	$0.50 / GPD	$0.51 / GPD Note 1	$0.48 / GPD Note 2	$0.48 / GPD Note 3
Initial End User Capital Cost For Membranes	$0.081 / GPD	$0.139 / GPD	$0.088 / GPD	$0.072 / GPD
Total Initial Capital Cost	$0.59/ GPD	$0.65 / GPD	$0.57 / GPD	$0.55 / GPD
Cost Comparison	**0%**	**10%**	**-3%**	**-6%**

- * Includes feed pumps, self-cleaning screen, equipment racks, backwash pumps and tanks, cleaning system, chemical feed systems.
- Note 1: Includes $4,000 additional cost due to having more membrane modules per skid charged at $2,000 USD per module.
- Note 2: Includes $44,000 cost savings due to having fewer membrane modules per skid charged at $2,000 USD per module. Includes $5,000 cost addition due to having higher pressure feed pump.
- Note 3: Includes $60,000 cost savings due to having fewer membrane modules per skid charged at $2,000 USD per module. Includes $9,000 cost addition due to having higher pressure feed pump.

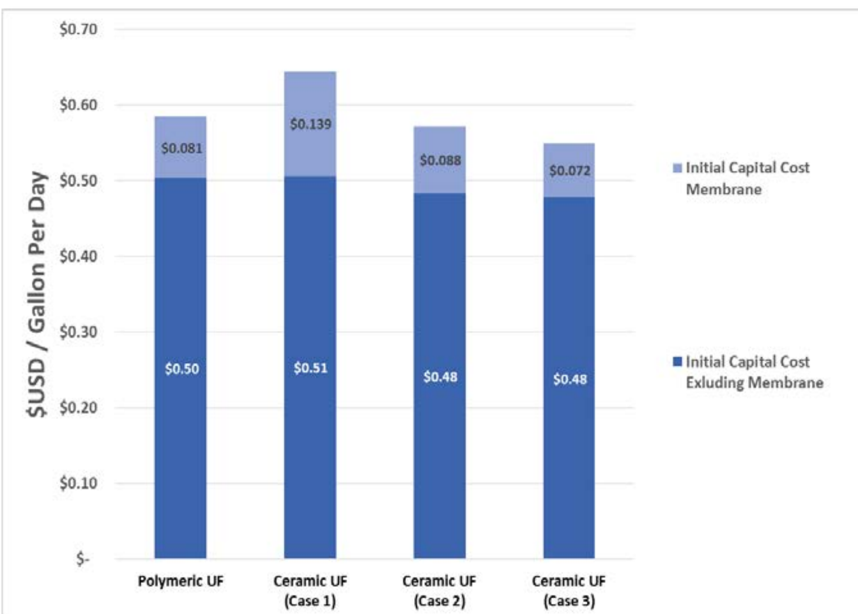

Figure 5: Graph of Initial Capital Cost Outputs of Techno-Economic Model Comparing Pressurized Polymeric UF Systems to Pressurized Ceramic UF Systems

Considering the benefits of the ceramic membrane in terms of the longer lifetime and more robust operation, the capital cost comparison on a system basis is competitive, even at the most conservative design flux case. With the higher flux cases, the savings in membrane modules far outweighs the cost of the higher pressure pump in terms of initial capital cost, and so there are cost savings up to 6% or $70,000 USD in this 2 MGD model.

The capital savings is attractive for the higher flux design cases, but there will be a penalty in terms of operating costs for the higher pressure needed as well as the additional coagulant chemical needed. In Table 4 below, the cost inputs for operating expenses for items such as power, coagulant, cleaning chemicals, and membrane replacement costs are listed as well as operating assumptions. These costs are then annualized, and 20 years of operating costs plus the initial capital costs are compared for a total cost of ownership analysis of the various cases.

For the chemical cleaning, it is assumed that all cases use the same levels of chemical cleaning including a maintenance cleaning with sodium hypochlorite and sodium hydroxide every 3 days, as well as an HCl acid maintenance cleaning every 9 days. A full recovery cleaning is assumed to take place every 3 months for all cases. The pilot data suggested that full recovery cleanings could take place on a frequency of 6 months to as long as a 1 year interval. However, for the operating cost evaluation, a more conservative recovery cleaning frequency is used.

For membrane replacement costs, the typical pressurized polymeric MF/UF membrane life is 5 to 7 years and the lifecycle model assumes a life span of 6 years. For ceramic UF membranes, the lifespan is estimated at 20 years. In this case, the polymeric MF/UF membrane price is also set at a typical value of $35USD/M^2 based on industry benchmarks. As noted earlier, the end user prices have a wide range, and so this price is a reasonable typical value for the purpose of this analysis.

For power consumption, the model uses a clean process water permeability value of 8 GFD/PSI (200 LMH/Bar) for the pressurized polymeric MF/UF membrane and 24 GFD/PSI (600 LMH/Bar) for the ceramic UF membrane. The operating pressure is then adjusted for flux and temperature and assumes an allowance for fouling to increase the TMP. Small pressure losses are also assumed for feed screen filters and piping. The backwash pump pressure is included in the power consumption as well as general power users such as the control system.

Table 4: Techno-Economic Operating Cost Model Inputs Comparing a Typical Polymeric MF/UF Case to a Ceramic UF System at Three Flux Level Cases

Model Inputs	Polymeric MF/UF System	Ceramic UF System Case 1	Ceramic UF System Case 2	Ceramic UF System Case 3
Flow Rate (Max)	2.0 MGD, 7,571 m^3/day			
Water Source	Direct River Water, Design 15°C (59°F) Temperature, < 10 NTU Turbidity			
Flux at max flow Rate	40 GFD (68 LMH)	116 GFD (196 LMH)	182 GFD (309 LMH)	225 GFD (381 LMH)
ACH Coagulant Dosage	1 mg/L	1 mg/L	2 mg/L	5 mg/L
Power Cost	$0.10 / kilowatt hour			
Chemical Costs	ACH Coagulant: $0.57/Kg, HCL: $0.50/Kg, NaOH: $0.10/Kg; NaOCl: $0.09/Kg			
Maintenance Cleaning	Every 3 days: 30 min. ambient cycle: 1,000 mg/L of NaOCl plus 600 mg/L of NaOH Every 9 Days: 30 min. ambient cycle: 1,000 mg/L of HCl			
Recovery Cleaning	Every 90 days: 4 hour heated cycle: 1,000 mg/L of NaOCl plus 600 mg/L of NaOH Every 90 days: 4 hour heated cycle: 1,000 mg/L of HCl			
Membrane Replacement Schedule	6 years		20 years	

Table 5: Techno-Economic Operating Cost Model Outputs Comparing a Typical Polymeric MF/UF Case to a Ceramic UF System at Three Flux Level Cases

Model Inputs	Polymeric MF/UF System	Ceramic UF System Case 1	Ceramic UF System Case 2	Ceramic UF System Case 3
Initial Capital Cost Including Membrane	$0.59 / GPD	$0.65 / GPD	$0.57 / GPD	$0.55 / GPD
20 Years Annualized Membrane Replace	$0.269 / GPD	$0.139 / GPD	$0.088 / GPD	$0.071 / GPD
20 Years Annualized Chemical Consumption	$0.089 / GPD	$0.091 / GPD	$0.114 / GPD	$0.199 / GPD
20 Years Annualized Power Consumption	$0.116 / GPD	$0.113 / GPD	$0.128 / GPD	$0.143 / GPD
Initial Capital Cost Plus 20 Years Operating Costs	$1.06 / GPD	$0.99 / GPD	$0.90 / GPD	$0.97 / GPD
Total Lifecycle Cost Comparison	**0%**	**- 7%**	**-15%**	**-9%**

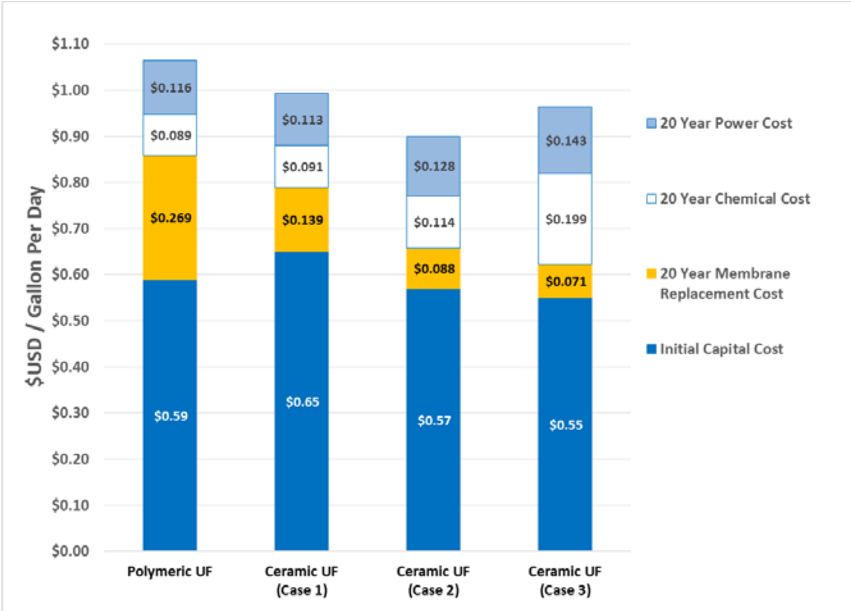

Figure 6: Graph of Initial Capital Cost and 20 Years of Operating Cost Comparing Pressurized Polymeric UF Systems to Pressurized Ceramic UF Systems

The total life cycle cost analysis above shows that in all cases, the ceramic UF membrane system has a lower total cost than the pressurized polymeric MF/UF system. The main cost savings is the membrane replacement costs, considering the 6 year life of a polymeric membrane versus a 20 year life of the ceramic membrane. With the higher flux cases for ceramic, there is a savings in capital cost and membrane replacement costs since there are fewer membranes, but there is an increase in power and chemical costs. This analysis shows that the most cost effective design for the ceramic membrane system is case two at a flux rate of 182 GFD and 2 mg/L of ACH coagulant dosage. This design is not the lowest in capital cost, but it is the optimum balance of initial capital cost and lifecycle cost. Comparing case two to the benchmark polymeric membrane case, the total savings in this example is $0.16 / GPD or $320,000 USD over a 20 year period. Considering case two is also 3% lower in initial capital cost compared to the polymeric membrane plant, should make a compelling case for the ceramic membrane alternative.

THE COLD WATER CASE

In many parts of the world, a surface water source will see low temperature periods during the 3-6 months of winter ranging from extremes of 0.3 to 5.0°C (32.5-41°F). For a polymeric MF/UF membrane plant operating in such cold conditions, the typical response is to reduce the flux during the cold periods. Where a plant would be designed for a flux rate of up to 45 GFD (76 LMH) at a temperature of 20°C (68°F), the same design would be reduced to just 28 GFD (48 LMH) if the temperature is reduced to 4°C (39°F), according to the design programs of most manufacturers. This leaves the designer with two choices: 1) design the plant to reach a design flow rate at the cold condition by adding more membrane area and reducing the flux; or 2) reducing the output of the plant during cold conditions. In most municipal drinking water plants, the typical choice is to reduce production or even shut down the plant during the winter months when water consumption is lower. In most industrial plants, the water demand is constant and so the flux of the polymeric membrane plant must be reduced to be able to keep up during the winter months.

There are several reasons why the polymeric membrane plant designs need to reduce their operating flux, as opposed to simply increasing the TMP to compensate for the increased water viscosity. One of the underlying reasons is that the polymeric membrane fibers constrict in extreme cold conditions, reducing the permeability beyond what the water viscosity change alone would predict. In a recent paper in the *Journal of Membrane Science*, a group of researchers from Lakehead University in Thunder Bay, Ontario, Canada discovered that in cold conditions there is a decrease in membrane lumen diameter, decrease in membrane permeability, an increase in the intrinsic hydraulic resistance, and an increase dextran rejection, suggesting that the pore size is constricting as well (Cui, 2017). They also noted that although most of the properties of the polymeric membranes returned to normal levels with increasing the water temperature, there was a permanent loss in permeability.

From an operations perspective, this cold water effect has long been observed by membrane plant operators. In a 2006 study funded by the AWWA Research Foundation (Pressdee, 2006), several membrane filtration plants were surveyed and in cases of cold water conditions the operators noted several changes required to the operations, including: 1) increased fiber breakage, 2) increased backwash durations, and 3) less effective chemical cleanings. In most cases sighted, the plant would reduce the operating flux and thereby reducing the TMP to lower the stress on the membrane fibers.

With ceramic membranes, however, there is a much lower coefficient of expansion due to temperature, and so there is no observed change in permeability beyond the change in water viscosity. This means that a plant can simply size the feed pump to run at a higher pressure during the cold periods without needing to reduce the plant output. Using the same techno-economic model as before, the capital and operating costs are compared but this time with an operating temperature of 4°C (39°F).

Table 6: Cold Water Condition Techno-Economic Capital Cost Model Comparing a Typical Polymeric MF/UF Case to a Ceramic UF System at Three Flux Level Cases

Model Inputs	Polymeric MF/UF System	Ceramic UF System Case 1	Ceramic UF System Case 2	Ceramic UF System Case 3
Flow Rate (Max)	2.0 MGD, 7,571 m^3/day			
Water Source	Direct River Water, Design 4°C (39°F) Temperature, < 10 NTU Turbidity			
Flux at max flow Rate	27 GFD (45 LMH)	116 GFD (196 LMH)	182 GFD (309 LMH)	225 GFD (381 LMH)
Total Membrane Area Total Module #	74K Ft^2 (7K m^2) 96 Modules	17K Ft^2 (2K m^2) 66 Modules	11K Ft^2 (1K m^2) 42 Modules	9K Ft^2 (1K m^2) 34 Modules
Initial Capital Cost Excluding Membrane*	$0.53 / GPD	$0.51 / GPD Note 1	$0.48 / GPD Note 2	$0.48 / GPD Note 3
Initial End User Capital Cost For Membranes	$0.121 / GPD	$0.139 / GPD	$0.088 / GPD	$0.072 / GPD
Total Initial Capital Cost	$0.66/ GPD	$0.65 / GPD	$0.57 / GPD	$0.55 / GPD
Cost Comparison	**0%**	**-2%**	**-13%**	**-16%**

- * Includes feed pumps, self-cleaning screen, equipment racks, backwash pumps and tanks, cleaning system, chemical feed systems.
- Note 1: Includes $60,000 cost savings due to having fewer membrane modules per skid charged at $2,000 USD per module. Includes $3,000 cost addition due to having higher pressure feed pump.
- Note 2: Includes $108,000 cost savings due to having fewer membrane modules per skid charged at $2,000 USD per module. Includes $8,000 cost addition due to having higher pressure feed pump.
- Note 3: Includes $124,000 cost savings due to having fewer membrane modules per skid charged at $2,000 USD per module. Includes $12,000 cost addition due to having higher pressure feed pump.

There is a clear advantage to the ceramic membrane plant designs that do not require a significant reduction in design flux to compensate for the cold water conditions. The added costs of more polymeric membranes into the benchmark design adds significant cost, such that the ceramic membrane plant costs are always lower in initial capital cost.

Applying the same operating cost inputs into the model for the cold water conditions shows even more total life cycle cost savings as compared to the warmer water cases above. The outputs of the design model here show savings up to 27%, and in this 2 MGD case, that represents $700,000 USD over 20 years.

Table 7: Cold Water Condition Techno-Economic Operating Cost Model Outputs Comparing a Typical Polymeric MF/UF Case to a Ceramic UF System at Three Flux Level Cases

Model Inputs	Polymeric MF/UF System	Ceramic UF System Case 1	Ceramic UF System Case 2	Ceramic UF System Case 3
Initial Capital Cost Including Membrane	$0.66 / GPD	$0.65 / GPD	$0.57 / GPD	$0.55 / GPD
20 Years Annualized Membrane Replace	$0.403 / GPD	$0.139 / GPD	$0.088 / GPD	$0.071 / GPD
20 Years Annualized Chemical Consumption	$0.098 / GPD	$0.091 / GPD	$0.114 / GPD	$0.199 / GPD
20 Years Annualized Power Consumption	$0.119 / GPD	$0.124 / GPD	$0.153 / GPD	$0.163 / GPD
Initial Capital Cost Plus 20 Years Operating Costs	$1.28 / GPD	$1.00 / GPD	$0.93 / GPD	$0.99 / GPD
Total Lifecycle Cost Comparison	**0%**	**- 22%**	**-27%**	**-23%**

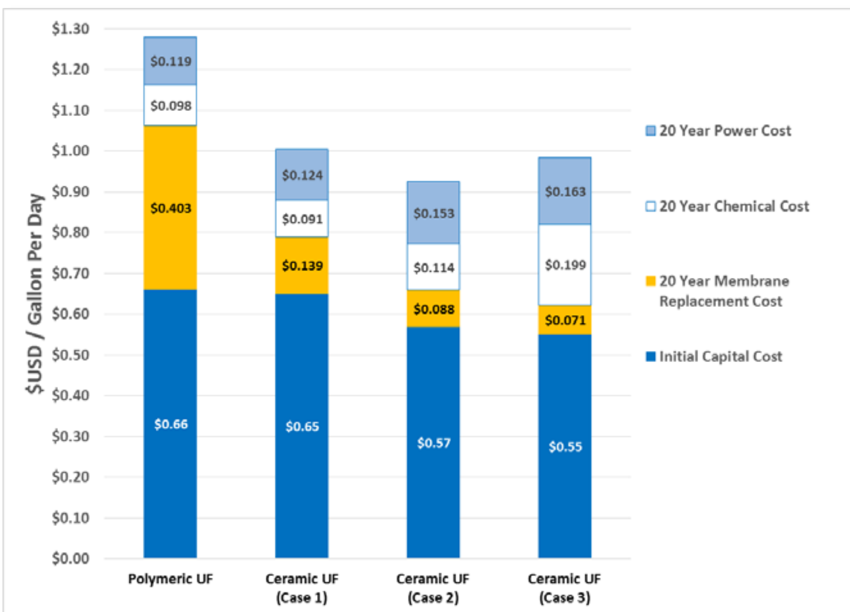

Figure 6: Graph of Cold Water Condition Initial Capital Cost and 20 Years of Operating Cost Comparing Pressurized Polymeric UF Systems to Pressurized Ceramic UF Systems

CONCLUSIONS

With the higher TSS, pressure, and chemical tolerance levels of the ceramic membrane, the system design can be tuned to fit the needs of the project with a much wider range than polymeric MF/UF membrane systems. Whether you have a feed water source with high turbidity events, or a need to increase the recovery, or a need to reduce initial capital cost, the ceramic membrane provides more flexibility to the designer and end user as compared to polymeric MF/UF membranes.

The advances made in production and process design of ceramic membranes have made them competitive with polymeric MF/UF membrane systems in the initial capital costs, in many cases. This is a significant change to conventional thinking where historically ceramic membrane systems where much more expensive. As shown in this analysis, the ceramic membrane system is competitive for a conservative design of even 100 GFD on a surface water source and moderate operating temperatures. And in cases of cold water operation, the ceramic membrane system has a significant capital cost savings over polymeric MF/UF membrane plants because the design flux can be maintained regardless of the change in temperature.

The analysis of operating costs shows the biggest contributor to overall life cycle costs are membrane replacements. With a 20 year expected life of the ceramic membrane, versus a typical 6 year life of the polymeric MF/UF membrane, there is a lower total life cycle cost in all design cases explored here. Considering the now competitive capital costs, the significant operational advantages, and the lower overall life cycle cost of ceramic membranes, they offer a compelling alternative to polymeric MF/UF membranes.

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