THE EVALUATION OF CLOTH MEDIA FILTRATION AS PRETREATMENT TO ULTRAFILTRATION IN WASTEWATER REUSE APPLICATIONS

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Abstract

As requirements for wastewater reuse quality are becoming increasingly more difficult across the United States, conventional tertiary filtration is often unable to deliver the required performance and reliability necessary to meet the new objectives. As a result, tertiary filtration is often supplemented or replaced with microfiltration (MF) or ultrafiltration (UF) membranes in the technology selection process. While some approaches couple the membranes with existing tertiary filtration systems, many applications directly apply clarified secondary effluent to the membranes. Unless the tertiary filters exist, a common perception holds that the required effluent quality can be achieved with membranes at a lower cost by eliminating the selection of intermediate tertiary filters. However, a strategy which employs tertiary filtration to pre-condition the water prior to membrane treatment offers distinct advantages, such as reduced operating costs, improved flexibility and increased reliability. Under most circumstances, the operation and maintenance (O&M) savings will offset the higher initial investment and yield a lower life cycle cost. To help determine the extent of the O&M reduction, this paper will compare operating parameters of a UF pilot system treating the same clarified secondary wastewater with and without the benefit of tertiary filtration.

Under the first treatment scenario, domestic wastewater is screened and treated with a conventional activated sludge system. The clarified water is then directly applied to the UF pilot without pre-conditioning by the tertiary filter. This approach eliminates equipment and O&M costs associated with the filter. However, the lack of pre-filtration results in lower permeability and reduced recovery through the UF membranes. In addition, this arrangement potentially increases the membrane area required to treat the same flow at the same pressure, which actually increases the overall O&M costs.

Under the second treatment scenario, the same clarified wastewater is sent through a full scale cloth media filtration (CMF) pilot system providing 5 m^2 of effective filtration area and featuring polyester cloth media. The filter uses an outside-to-inside, vertically mounted configuration that allows heavier solids to settle to the bottom of the filter tank, where they are periodically vacuumed to drain. As a filtration layer develops on the cloth, the tank level rises until a high setpoint is reached, which initiates a backwash step. The high quality filtered water that passes through the cloth represents the influent to the subsequent UF treatment system.

The UF system used in the evaluation of both scenarios consists of a single vertical 60 m^2 module provided by inge GmbH, a subsidiary of BASF. The module contains multibore fibers constructed of polyethersulfone (PES) that filter the wastewater using an inside-to-outside flow path that eliminates the need for cleaning with an air-scour system. Wastewater flows alternately into the top and bottom of the fibers in order to achieve an even layer of solids on the membrane surface, which improves the effectiveness of the periodic backwash events. The UF system integrates a flow-paced coagulant feed system, which can be set to inject metal salts into the feed water to optimize permeability. To maintain low trans-membrane pressures (TMP), the backwash flow is routinely injected with one of three cleaning chemicals: caustic for removal of organic foulants, hydrochloric acid to control inorganic fouling, or sodium hypochlorite for disinfection.

Each scenario consists of duplicate pilot runs to benchmark chemical dosage, flux, TMP, permeability and effluent quality to establish the optimum system O&M requirements. This paper will compare and contrast the operating data acquired during the pilot runs in order to assist in the life-cycle evaluation of each scenario.

Materials and Methods

Experimental Facilities

The CMF and UF pilot systems used in this research are located at the Rock River Water Reclamation District (RRWRD) in Rockford, IL. The facility consists of a conventional sewage treatment system; i.e., mechanical screening and sedimentation followed by activated sludge oxidation and secondary sedimentation. Then, the secondary settled effluent (SSE) is disinfected with sodium hypochlorite and discharged to the adjacent Rock River.

In the first experimental scenario (without tertiary filter), a portion of the SSE was fed through the UF pilot plant. A submersible pump in the weir feeding the chlorine contact basin pumped SSE through a 300 μm wire strainer, which was used to remove large particles that could damage the membrane fibers. Once through the strainer, the flow filled a 120-gallon feed tank.

Figure 1 shows the flow diagram for the UF pilot plant. During filtration modes, SSE was pumped from the feed tank through a single 60 m2 UF module using a 3 HP feed pump. The module used was inge's dizzer® XL and contained about 2,000 Multibore® fibers, each with (7) 0.9 mm bores (lumen). The size of the membrane pores on the inside of each bore was 0.02 μm (nominal), resulting in a molecular weight cutoff of 200,000 Daltons and more than 4 log virus removal.

Aluminum sulfate (alum) was injected into the SSE prior to entering the membrane module in order to coagulate colloidal material into larger particles, which minimized plugging of the membrane pores and, therefore, resulted in higher flows through the membrane without increasing feed pressure (i.e., higher permeabilities). The UF pilot plant also included a piping loop between the alum injection point and the membrane module that was made up of five separate chambers designed to allow adjustment of the alum retention time; however, all five chambers were used for each of the pilot runs.

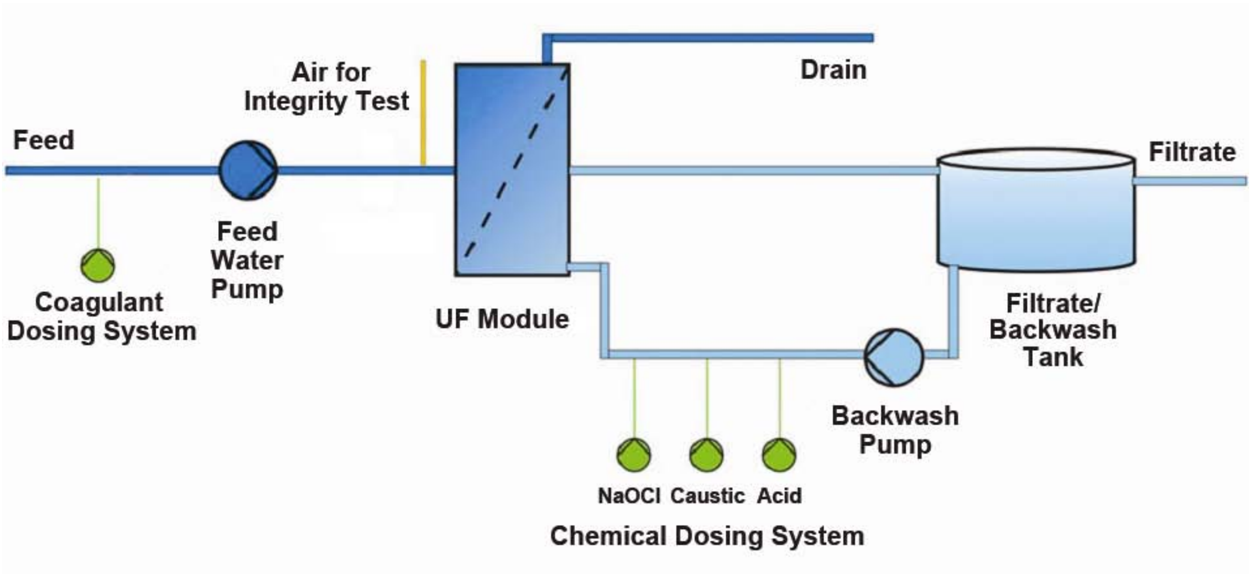

Figure 1. Flow Diagram for the UF Pilot Plant

During filtration, SSE was introduced to the inside of each fiber bore with filtrate (permeate) and forced to the outside of the membrane (i.e. inside-out). The filtrate flowed to the filtrate/backwash tank, where it overflowed into the effluent line and was discharged into the chlorine contact tank.

A variable-frequency drive (VFD) on the membrane feed pump was used to maintain a constant flow through the membrane during each trial. The feed pressure to the membrane was monitored, and was found to increase throughout the filtration process in relation to the degree of fouling. To remove foulants from the membrane surfaces, a periodic backwash was performed in which filtrate was pumped in the reverse direction through the membrane (from outside the fibers to inside the fiber bores) using a 5 HP backwash pump. The filtration interval between backwashes was varied during some of the trials to observe its effect on the system O&M requirements; some trials used a 30-minute filtration interval, while other trials used a 45-minute interval.

To maintain higher permeabilities, the membrane was cleaned periodically with caustic, hydrochloric acid, and/or sodium hypochlorite. This was accomplished by dosing the chemical into the backwash water for approximately 80 seconds, soaking the membrane in the chemical for 10 - 15 minutes, and then flushing it to drain with filtrate. The frequency of these chemically-enhanced backwashes (CEBs) was varied during several of the trials to see what affect this had on O&M requirements.

Two sampling points were used during this experiment to collect the wastewater samples. The first sampling point was located on the feed line prior to the alum injection point, and the second was located on the filtrate line immediately after the module. In addition, the pilot plant included the following instruments for monitoring parameters via the pilot's human-machine interface (HMI):

- feed and filtrate/backwash pressure transmitters,
- feed temperature sensor,
- feed and filtrate/backwash flow meters,
- feed and filtrate/backwash tank level sensors,
- filtrate/backwash pH meter, and
- common feed and filtrate/backwash turbidimeter (with sample pump and valves).

In the second experimental scenario (with tertiary filter), a portion of the SSE was fed through the CMF pilot unit. A submersible pump in the weir feeding the chlorine contact basin pumped SSE into the unit. As shown in Figure 2, this feed flowed over the influent weir, through the unit's single 5 m^2 disk, into the centertube, over the effluent weir, and into the discharge line. The outside of the disk was wrapped with OptiFiberPES-14® polyester microfiber cloth media, with a particle sized removal rating of 5 μm (nominal).

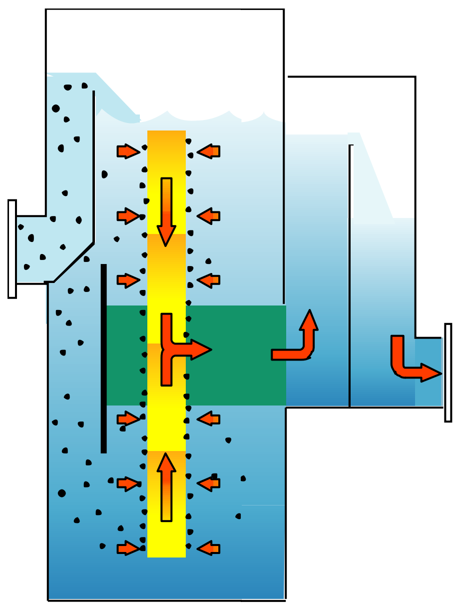

Figure 2. Filtration Flow Diagram for the CMF Pilot Plant

As the amount of particulates on and within the cloth media increased, the pressure required to drive water through the media increased. This resulted in an increased water level within the filter basin and increased differential pressure on the media. Upon reaching a specific basin water level, the unit went into backwash, shown in Figure 3. During backwash, the unit's drive motor rotated the center tube while a centrifugal pump pulled filtered water from inside the disk through a suction header pressed against the outside of the filter cloth. This high-velocity reverse flow drew the particulates from the cloth media, renewing the cleaning capacity and lowering the basin water level and media differential pressure.

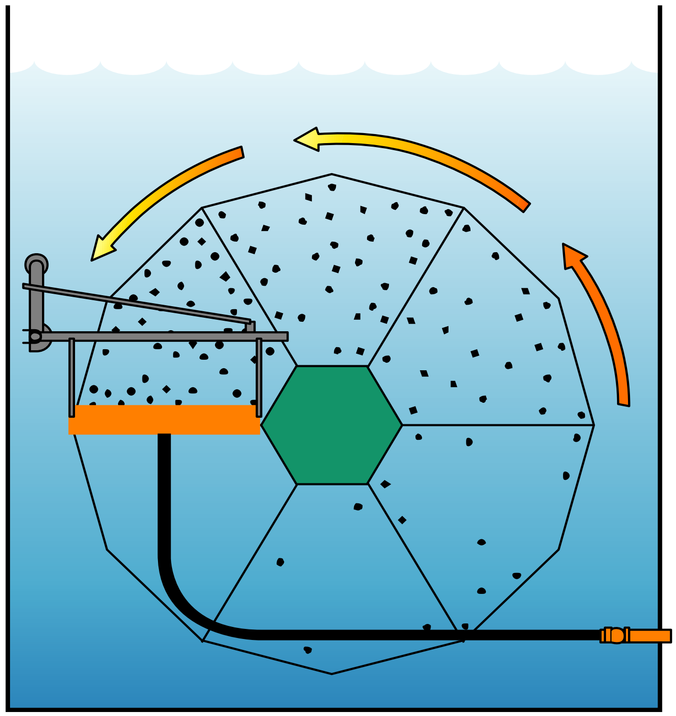

Figure 3. Backwash Flow Diagram for the CMF Pilot Plant

The filter's outside-in flow path allowed heavier particulates to settle to the bottom of the tank, as shown in Figure 4. Periodically, a pump and associated piping removed the accumulated sludge from the tank. The system was set to remove the settled sludge every 6 backwashes for all pilot trials.

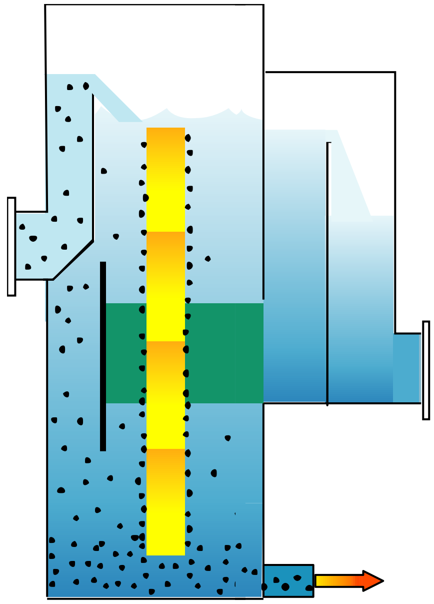

Figure 4. Sludge Removal Flow Diagram for the CMF Pilot Plant

From the CMF pilot, the flow moved by gravity to the UF pilot system, filling up its 120-gallon feed tank. The flow through the UF pilot plant was as described earlier for the first experimental scenario. The CMF pilot (left) and UF pilot (right) are shown in Figure 5.

Figure 5. CMF and UF pilots within the Research Center at the RRWRD

In addition to the two sampling points on the UF pilot plant, a sample point on the CMF influent was used to check feed water quality.

Key Parameters

One of the key parameters calculated during the experiment was membrane flux (MF). This is the rate at which flow (Q) passes through a given membrane area (A) and is calculated using the following equation:

$$MF = Q / A$$

The flux is defined in gallons per square foot of membrane per day (gfd) or liters per square meter of membrane per hour (lmh). In all of the trial runs of the pilot systems, the feed flow was set to a specific value, and the speed of the membrane feed pump was automatically adjusted to maintain this value on the feed flow meter. Since the membrane area remained constant at 60 m^2, the flux remained constant during each trial.

There are also two types of fluxes to be concerned with: gross flux (MF_G) and net flux (MF_N). Gross flux is the maximum flow needed to achieve the required effluent flow (Q_E), while net flux is based solely on the effluent flow. Gross flux takes into consideration how much time the system is actually in filtration mode and how much of the effluent produced is used for backwashes and CEBs. The equation for each flux is as follows:

$$MF_N = Q_E / A$$

$$MF_G = MF_N \; x \; (1440 \,/\, total\ daily\ filtration\ time) \; x \; (Q_E + Q_B) / Q_E)$$

where Q_B is the daily volume of filtrate used for backwashes and CEBs. Since MF_G often determines membrane quantity, a higher MF_G means that more membranes will be needed to get the required Q_E out of the system.

Another key parameter calculated during the experiment was trans-membrane pressure (TMP). This is the pressure drop the flow undergoes as it goes through the membrane and is calculated by determining the difference between the feed pressure (P_{feed}) and the filtrate pressure ($P_{filtrate}$):

$$TMP = P_{feed} - P_{filtrate}$$

The TMP is expressed in terms of psi or bar. During membrane filtration at a constant flux, the TMP will gradually rise as the membrane slowly accumulates surface solids and foulants; therefore, a periodic backwash or chemical cleaning is needed to keep the TMP low and maintain the desired flux (flow). It's also important to note that feed temperature significantly affects system TMP; the colder the feed water, the higher the water density and viscosity, which increases the TMP.

The most important parameter for determining the O&M differences between the two experimental scenarios is the membrane permeability (MP). This is the gross flux required to achieve a certain TMP and is calculated as follows:

$$MP = MF_G / TMP$$

One important observation that can be made from this equation is how permeability will change with a change in flux. As a basic rule of thumb, doubling the flow through an opening will quadruple the pressure through that opening [Lau (2008)]; therefore, as the gross flux MF_G increases, the TMP will increase much more, resulting in a lower membrane permeability MP.

Permeability is typically expressed in gfd per psi (gfd-p) or lmh per bar (lmh-b). The system that runs at a higher permeability will often be able to use fewer membranes to achieve the same flow at the same TMP, which reduces the O&M requirements by lowering membrane replacement costs, backwash pump energy, and chemical usage. Since flux is temperature-sensitive, it's critical that all permeabilities be normalized to the same temperature before comparing permeabilities from different pilot runs. Therefore, all permeabilities calculated during these experiments were normalized to 20°C using the following formula:

$$MP_N = MP_T / K$$

Where MP_N is the membrane permeability normalized to 20°C, MP_T is the membrane permeability at temperature T, and K is the temperature correction factor shown in Figure 6.

Another key parameter to use in evaluating membrane O&M requirements is system recovery (SR). This is the percentage of total influent flow (Q_I) that actually becomes effluent flow (Q_E) and is calculated using the following equation:

$$SR = (Q_E / Q_I) \times 100\%$$

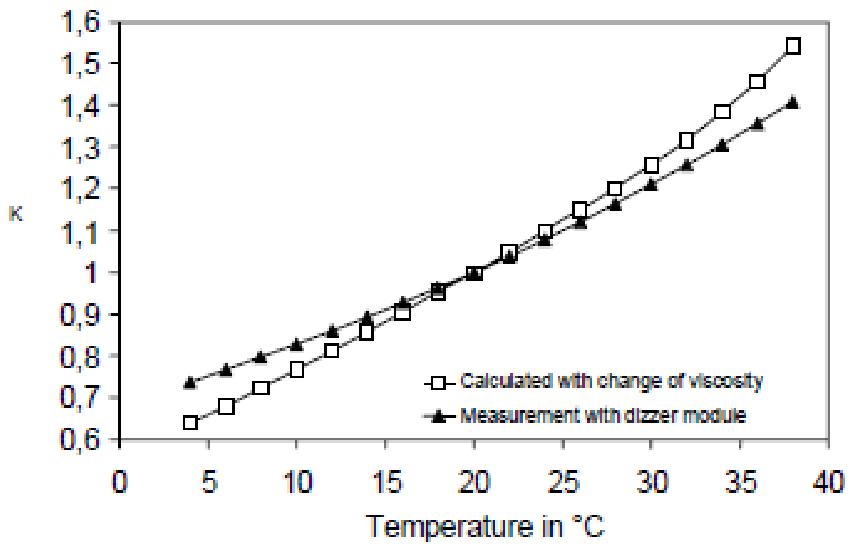

Figure 6. Temperature Correction Factor (K)

The influent and effluent flows are typically expressed in gallons per day (gpd) or cubic meters per day (m^3/d). The difference between Q_I and Q_E is the volume of filtrate that is used for backwash and chemical dilution; the more frequent the backwashes or CEBs, the lower the recovery. A lower SR means that more membranes will be needed to get the required Q_E out of the system.

Experimental Design

The entire two-phase experiment was conducted from February 2012 through October 2012. Based on both previous pilot experience and inge's recommendations [Koti (2011)], several of the UF pilot parameters remained unchanged throughout the entire experiment, as shown in Table 1.

The operating sequences of the UF pilot were the same for both phases, with the exception of the two Forward Flush steps, which were not used during the last part of the first phase (without tertiary filter) or during the entire second phase (with filter). The filtration sequence consisted of four steps: filtration bottom, backwash bottom, filtration top, and backwash top. By alternating the direction of filtration flow, feed particles were more evenly distributed along the length of each fiber bore, resulting in a more even and efficient backwash flow. The backwash flow was alternated in the same fashion to make sure the solids collected in the open area at both ends of the module were flushed to drain.

During four of the five phase 1 time periods, two additional steps were used during the filtration sequence: forward flush bottom and forward flush top. These steps consisted of a brief crossflow through the bores following a backwash to remove any solids not removed by the backwash. The steps were discontinued because they were found to have little to no effect on the system permeability.

Table 1. UF Pilot Parameters That Remained Constant Throughout the Experiment

Description	Units	Value
Coagulation Chambers Used	No.	5
Backwash Bottom (BWB)	Sec.	60
Backwash Top (BWT)	Sec.	60
CEB Backwash Top	Sec.	30
CEB Backwash Bottom	Sec.	30
Chemical Introduction Top	Sec.	40
Chemical Introduction Bottom	Sec.	20
CEB1 Soak Time	Min.	15
CEB2 Soak Time	Min.	15
CEB3 Soak Time	Min.	10
CEB Rinse Top	Sec.	40
CEB Rinse Bottom	Sec.	40
Bulk Alum Concentration	%	48.5
Bulk NaOH Concentration	%	50
Bulk HCl Concentration	%	32
Bulk NaOCl Concentration	%	6
Caustic Introduction pH	pH	12.2
Acid Introduction pH	pH	2
NaOCl Introduction Concentration	mg/l Cl_2	20

The CEB sequence consisted of seven steps for each of the three chemicals used: backwash top, backwash bottom, chemical introduction top, chemical introduction bottom, soak, rinse top, and rinse bottom. The backwash steps were identical to those used during the filtration sequence except for only half the time; they were used to remove as many solids as possible, minimizing the amount of chemical needed. The chemical introduction and soak steps allowed the membranes to contact and soak in the chemical solution, which dissolved nearly all of the solids that the backwash hadn't removed. The rinse steps were done to flush the chemical out of the module and to the drain.

The CEB1 sequence involved cleaning with caustic first, followed by cleaning with hydrochloric acid; a 5-minute filtration bottom step was placed between the two cleaning sequences to ensure that all of the caustic was flushed out of the module before the introduction of acid to avoid any issues with isolated overheating. The CEB2 sequence involved cleaning with hydrochloric acid only, and the CEB3 sequence involved cleaning/disinfecting with sodium hypochlorite.

Typically, the CEB sequence will return the permeability to the level attained at the start of the previous filtration sequence. However, there are certain atypical circumstances that can cause this not to occur: prolonged shutdown without flushing, loss of alum or cleaning chemicals, prolonged power failure, etc. This happened several times during the course of the pilot study, requiring that the UF system be put through an intensive clean-in-place (CIP) sequence, whereby the normal cleaning chemicals are heated and recycled one-by-one through the membrane fibers in a crossflow pattern, then allowed to soak the membranes for a prolonged period of time. The chemical is then flushed out to drain with filtrate.

The first experimental scenario (without tertiary filter) was begun in February 2012 and completed in June 2012. This phase of the pilot was divided into five separate time periods in which several different parameters were changed. The permeability equation given earlier shows that lower fluxes yield higher permeabilities; however, higher fluxes typically result in lower O&M costs because membrane replacement normally accounts for the largest portion. In addition, the higher feed pump energy usage required to obtain the higher fluxes is often offset by the savings in backwash pump power and chemical usage required with fewer membranes [Zheng (2011)]. For these reasons, the goal of the first experimental scenario was to adjust various parameters to determine the combination that would give the highest sustainable permeability at the highest net flux.

During the first phase, gross fluxes were varied from 40 to 78 lmh by adjusting the feed flows from 10.6 to 20.6 gpm. The UF pilot was operated at each flux for several days, and the system flows and pressures were measured and recorded on both the HMI and the unit's personal computer (PC). The filtration sequence parameters that were adjusted are given in Table 2.

Table 2. Variable Filtration Sequence Settings for Phase 1 (without filter)

Parameter	Units	2/21-3/14	3/14-3/22	3/22-3/25	3/25-4/18	4/18-6/26
Filtration Bottom	Min.	45	30	45	30	30
Forward Flush Bottom	Sec.	30	30	30	30	--
Filtration Top	Min.	45	30	45	30	30
Forward Flush Top	Sec.	30	30	30	30	--

Likewise, some of the chemical settings were adjusted during the first phase, as shown in Table 3. A CEB 2 interval of 999 hours was set so that this cleaning regimen (acid only) wouldn't occur during that particular pilot run without a previous cleaning with caustic. During the last time period, alum was injected for the first 10 minutes of each filtration bottom and filtration top step, in lieu of continuously; this was to determine if less alum could be used to achieve the same permeability.

A CIP sequence was performed two times during the first phase: on March 6-8, 2012 following a prolonged shutdown to fix leaks, and again on May 8-10, 2012 following an unusually high flux (up to 75 lmh) that caused low permeability (< 100 lmh-b) and premature membrane fouling. Following each CIP, the permeability returned close to a clean-membrane level.

Table 3. Variable Chemical Settings for Phase 1 (without filter)

Parameter	Units	2/21-3/22	3/22-4/12	4/14-6/26
CEB 1 Interval	Hours	12	8	6
CEB 2 Interval	Hours	999	12	15
CEB 3 Interval	Hours	8	36	48
Alum Dosage	mg/L Al^{+3}	1.5	3.0	5.0
Alum Injection Duration	Type	Continuous	Continuous	Pre-Coat

The second experimental scenario (with tertiary filter) was begun in August 2012 and completed in October 2012. This phase of the pilot was divided into ten separate trials in which two main parameters were changed: gross flux and filtration interval. As in phase 1, the goal of this phase was to determine the combination that would give the highest sustainable permeability at the highest net flux. However, the results from phase 1 were used to set several of the parameters for phase 2, as shown in Table 4. These are in addition to those shown earlier in Table 1.

Table 4. UF Pilot Parameters That Remained Constant Throughout Phase 2

Description	Units	Value
Forward Flush Bottom	Sec.	0
Forward Flush Top	Sec.	0
CEB 1 Interval	Hours	12
CEB 2 Interval	Hours	999
CEB 3 Interval	Hours	72
Alum Dosage	mg/L Al^{+3}	2
Alum Injection Duration	Type	Continuous

During the second phase, gross fluxes were varied from 45 to 85 lmh by adjusting the feed flows from 11.9 to 22.5 gpm. The UF pilot was operated at each flux for several days, and the system flows and pressures were measured and recorded on both the HMI and the unit's personal computer (PC). The UF parameters that were adjusted are given in Table 5.

A CIP sequence was performed four times during the second phase: on August 20-22, 2012 following a prolonged power outage, on September 3-5, 2012 following a prolonged shutdown, on October 4-5, 2012 following a prolonged power outage, and again on October 18-19, 2012 following another prolonged power outage. Following each CIP, the permeability returned close to a clean-membrane level.

Table 5. Variable UF Settings for Phase 2 (with filter)

Trial Number	Gross Flux (lmh)	Feed Flow (gpm)	Filtration Bottom (min.)	Filtration Top (min.)
1	50	13.2	30	30
2	60	15.9	30	30
3	70	18.5	30	30
4	70	18.5	45	45
5	75	19.8	30	30
6	75	19.8	45	45
7	80	21.1	30	30
8	80	21.1	45	45
9	85	22.5	30	30
10	85	22.5	45	45

Results and Discussion

Feed Water Characteristics

Table 6 lists the analyzed and observed parameters of the SSE feed to the UF pilot during phase 1 (no tertiary filter), and Table 7 lists the SSE characteristics during phase 2 (with tertiary filter).

Table 6. Characteristics of the Influent to the UF Pilot During Phase 1 (June 2012)

Parameter	Unit	Min	Max	Average
pH	pH	7.0	7.2	7.1
Temperature	°F	63	69	66
Turbidity	NTU[1]	1.1	2.2	1.3
Total Suspended Solids (TSS)	mg/L	1	7	2.7
Chemical Oxygen Demand (COD)	mg/L	14	38	22.3
5-Day Biological Oxygen Demand (BOD_5)	mg/l	3	9	5.4
Carbonaceous BOD (CBOD)	mg/l	2	7	3.5

[1]Nephelometric Turbidity Unit (NTU) – these values are estimates

Table 7. Characteristics of the Influent to the Tertiary Filter During Phase 2 (Aug-Sep 2012)

Parameter	Unit	Min	Max	Average
pH	pH	6.8	7.0	6.9
Temperature	°F	68	72	70
Turbidity	NTU[1]	1.62	6.38	4.13
Total Suspended Solids (TSS)	mg/L	1	20	8.5
Chemical Oxygen Demand (COD)	mg/L	24	49	34.2
5-Day Biological Oxygen Demand (BOD_5)	mg/l	5	16	9.7
Carbonaceous BOD (CBOD)	mg/l	3	10	5.6

[1]Nephelometric Turbidity Unit (NTU)

UF Pilot Performance During Phase 1 (no tertiary filter)
Data was collected from the UF pilot operating at the different gross fluxes to see how membrane performance was affected. Figure 7 shows the normalized membrane permeabilities (MP_N) of two separate runs operating at different fluxes (55 and 78 lmh).

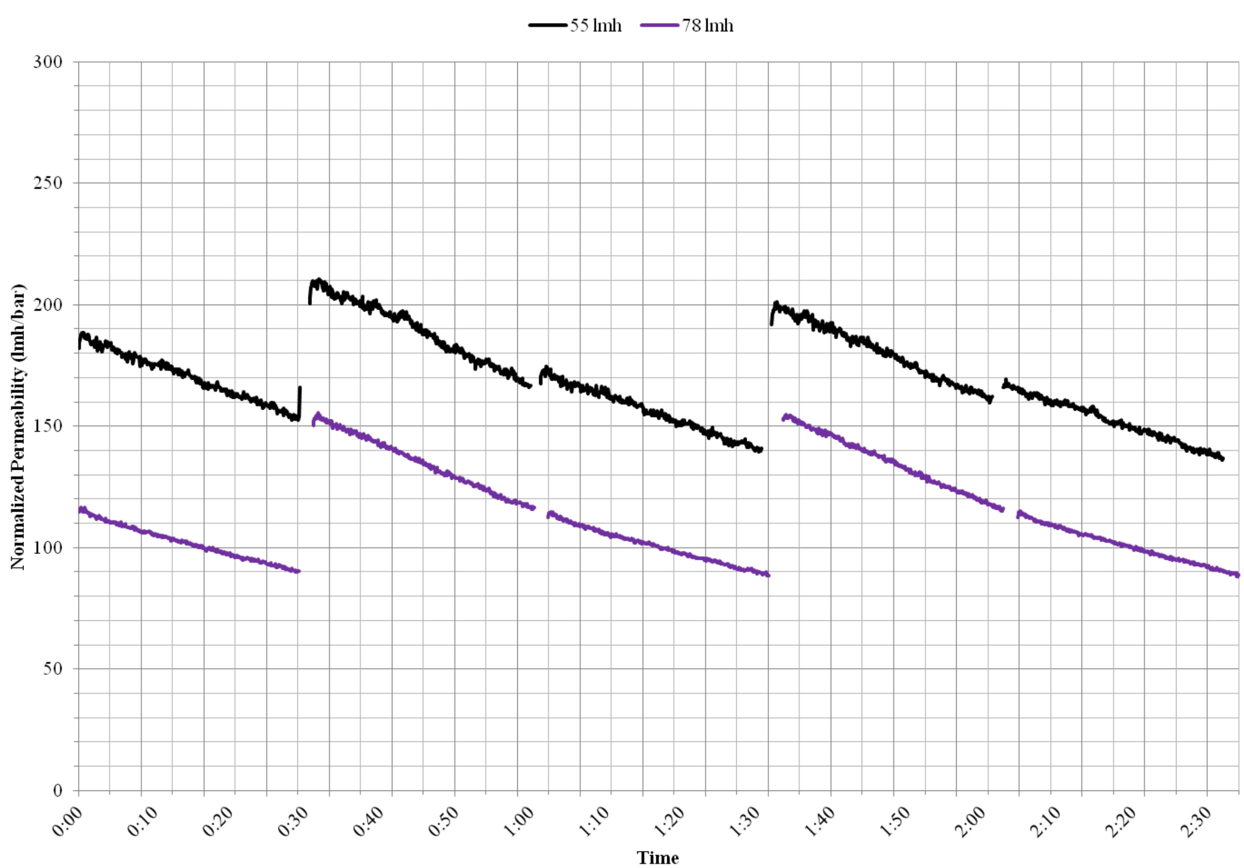

Figure 7. Normalized Permeabilities at Two Gross Fluxes for Phase 1 (without filter)

As you can see, the higher flux shows a marked decrease in permeability. As noted earlier, the decreasing permeability will add to the O&M costs while the increasing flux will reduce the

O&M costs. It was also noted earlier that this will typically result in a net reduction in O&M costs. For instance, the average normalized permeability decreases from about 170 to 115 lmh/bar when the flux is increased from 55 to 78 lmh. The resultant energy increase can be calculated using the formula: kwh/yr = flow x head increase x 0.746 x 24 hrs x 365 days/ 3960 / pump efficiency.

For a 1 mgd (694 gpm) UF plant and a feed pump head increase and efficiency of 5 ft and 60%, respectively, the annual energy required will be 9,554 kwh. Assuming an energy cost of $0.08/kwh, the additional annual power cost will be $764.

On the other hand, increasing the membrane flux by 42% will reduce the membrane quantity by that same percentage, along with the plant O&M costs. The total O&M costs for a UF plant are estimated as 19,000 € (25,000 USD) per mgd [Zheng (2011)]; therefore, the resultant O&M savings will be $10,500/yr. Subtracting the additional feed pump power cost of $764 yields a net annual cost savings of $9,736.

UF Pilot Performance During Phase 2 (with tertiary filter)

During this phase, the CMF effluent was monitored to determine the improvement of the membrane feed quality under this scenario. Table 8 shows the CMF effluent characteristics.

Table 8. Characteristics of the Cloth Media Filter (CMF) Effluent During Phase 2

Parameter	Unit	Min	Max	Average
Turbidity	NTU[1]	1.25	2.14	1.69
Total Suspended Solids (TSS)	mg/L	ND[2]	8.6	1.3
Chemical Oxygen Demand (COD)	mg/L	11	22	18

[1]Nephelometric Turbidity Unit (NTU)
[2]Non-Detectable

Comparing Tables 7 and 8, the CMF has removed an average of 59, 85, and 47% of the influent turbidity, TSS, and COD, respectively.

As was done during phase 1 testing, the UF pilot was operating at different gross fluxes to see how membrane performance was affected. Figure 8 shows the normalized membrane permeabilities (MP_N) of three separate runs operating at different fluxes (45, 60, and 70 lmh).

As would be expected, there's a decrease in permeability with each increase in flux, but it's far less pronounced than recorded in phase 1 (without filter). In addition, nearly all of the permeabilities are higher than those shown in Figure 7 for phase 1. This is due mostly to the fact that the TSS to the UF system is 85% less when the CMF is included.

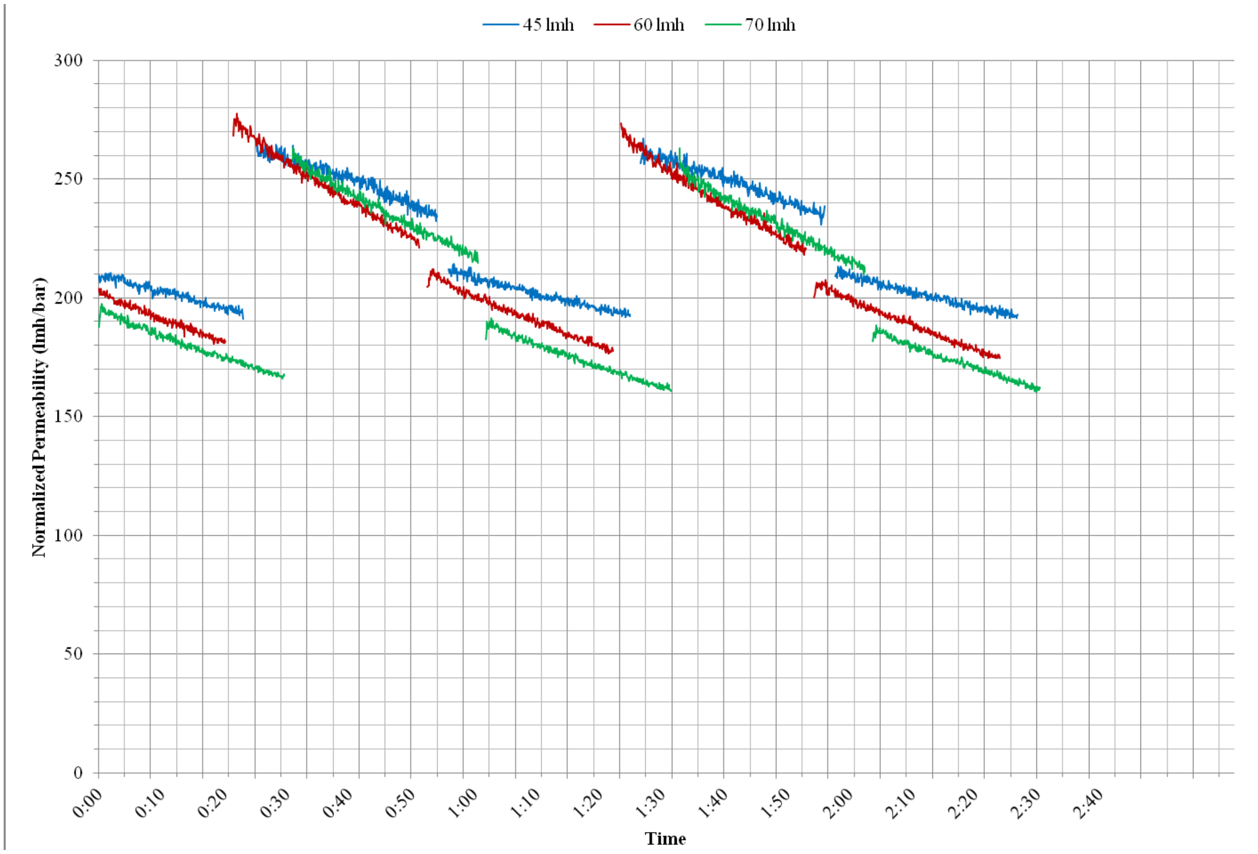

Figure 8. Normalized Permeabilities at Varying Gross Fluxes for Phase 2 (with filter)

Comparison of Phase 1 and 2 UF Pilot Performance

The above observation is confirmed in Figure 9, where the average normalized permeabilities are plotted for varying gross fluxes for both phases, with and without the tertiary filter. Since the higher fluxes will have lower O&M costs, the permeability differential at the greater fluxes is of most significance. For instance, comparing the data at 75 lmh gross flux shows us that the permeability for phase 2 (with filter) is about 30% higher than the permeability for phase 1. Given that the annual O&M cost for a UF plant is about $25,000 per mgd, this will result in an annual O&M savings of 30% of this, or $7,500 per mgd.

When looking at the 20-year life cycle cost for both phases, it's also necessary to consider the capital cost for the CMF and UF system as well as the O&M costs. The cost for a membrane treatment system is estimated to be 2000 € per m^3/hr (0.42 USD per gpd) [Zheng (2011)]; therefore, a 30% capacity increase for a 1 mgd plant will be a $0.14 per gpd reduction, which is very close to the capital cost for the 1 mgd CMF system. The O&M cost for the CMF system will be very small, estimated at $1,000 per mgd.

The result is that the UF capital savings offsets the CMF capital cost, and the actual savings will be the net O&M reduction, which is $7,500/yr for the UF less $1,000/yr for the CMF, or $6,500/yr. Over the 20-year life of the plant, this comes out to a total savings of $130,000, not including the interest. The building needed to house the equipment has not been considered as it's assumed the footprint increase for the CMF will be roughly the same as the footprint decrease for the smaller UF system.

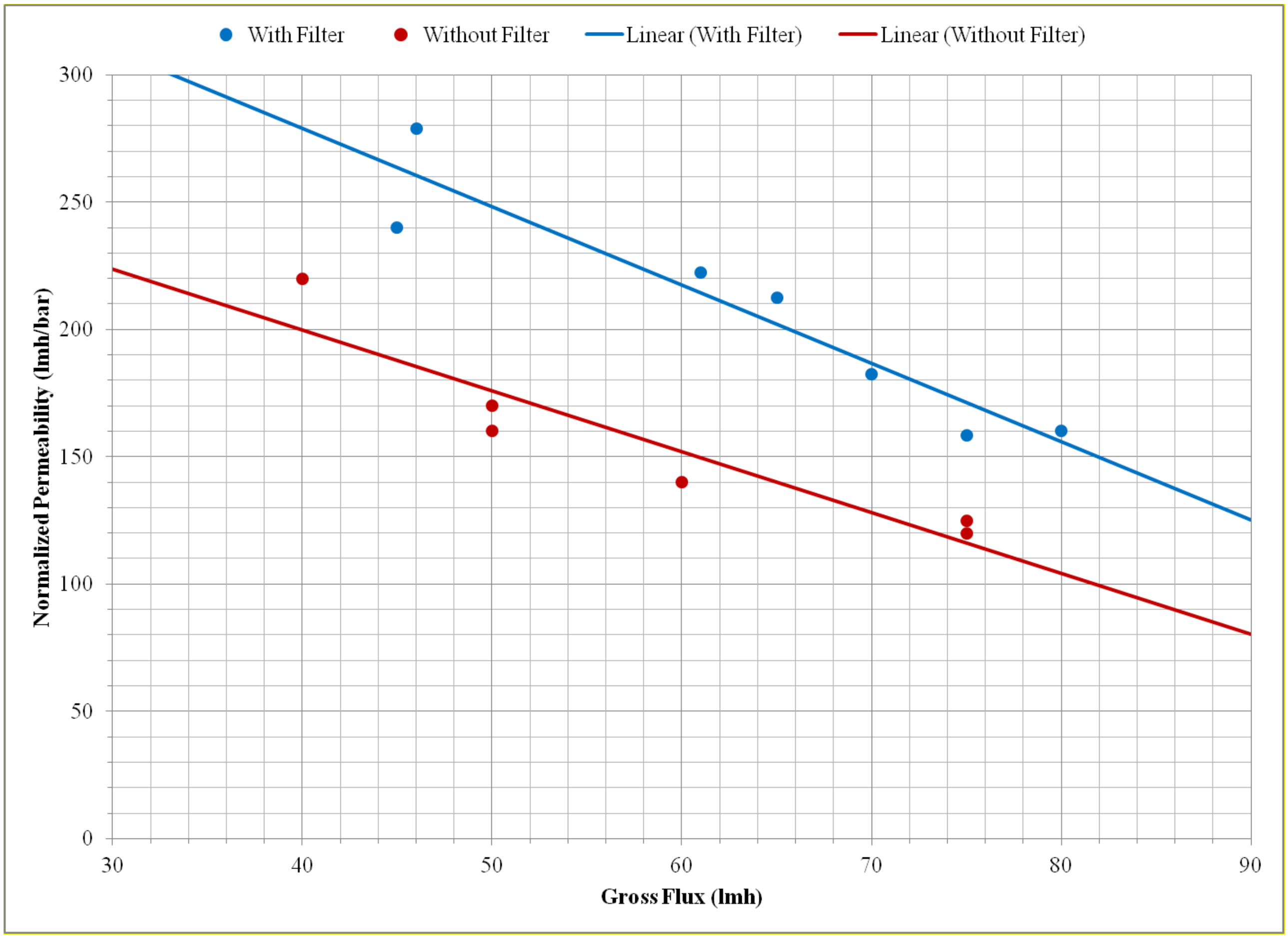

Figure 9. Comparison of Normalized Permeabilities for Both Phases (with and without filter)

In addition to the permeability comparisons between the two phases, two other observations were made that will affect the O&M costs for the treatment system. First, there were several instances where the upstream secondary system sent slugs of high solids effluent (> 30 mg/l TSS) to the CMF or UF system. When this occurred during phase 1 of the experiment (without tertiary filter), the UF operating parameters had to be immediately adjusted in order to keep from overloaded the membrane with solids. The UF system controls can be programmed to make these adjustments automatically, but the risk of overloading the membrane is much higher in this scenario than if a tertiary filter is there to handle the higher solids loading. In addition, more frequent CIP cleanings are inevitable. When these solids spikes occurred during phase 2 (with the tertiary filter), the filter was able to handle the extra loading with temporary increases in backwash frequency, and the downstream membranes saw only a minor increase in their influent turbidity. To illustrate this, Figure 10 gives the influent and effluent turbidity for the CMF pilot during a 5-day period in October 2012.

Second, there were several times during phase 2 operation when the CMF effluent turbidity was around 1 NTU. Depending on the final effluent requirement, it's highly possible that there will be situations where the CMF is capable of meeting the effluent requirements without the membranes or by sending only a portion of the flow through the membranes and blending the filtrate back with the CMF effluent. If this is the case, the UF system may not have to be operated all the time at full capacity, which could result in significant O&M savings.

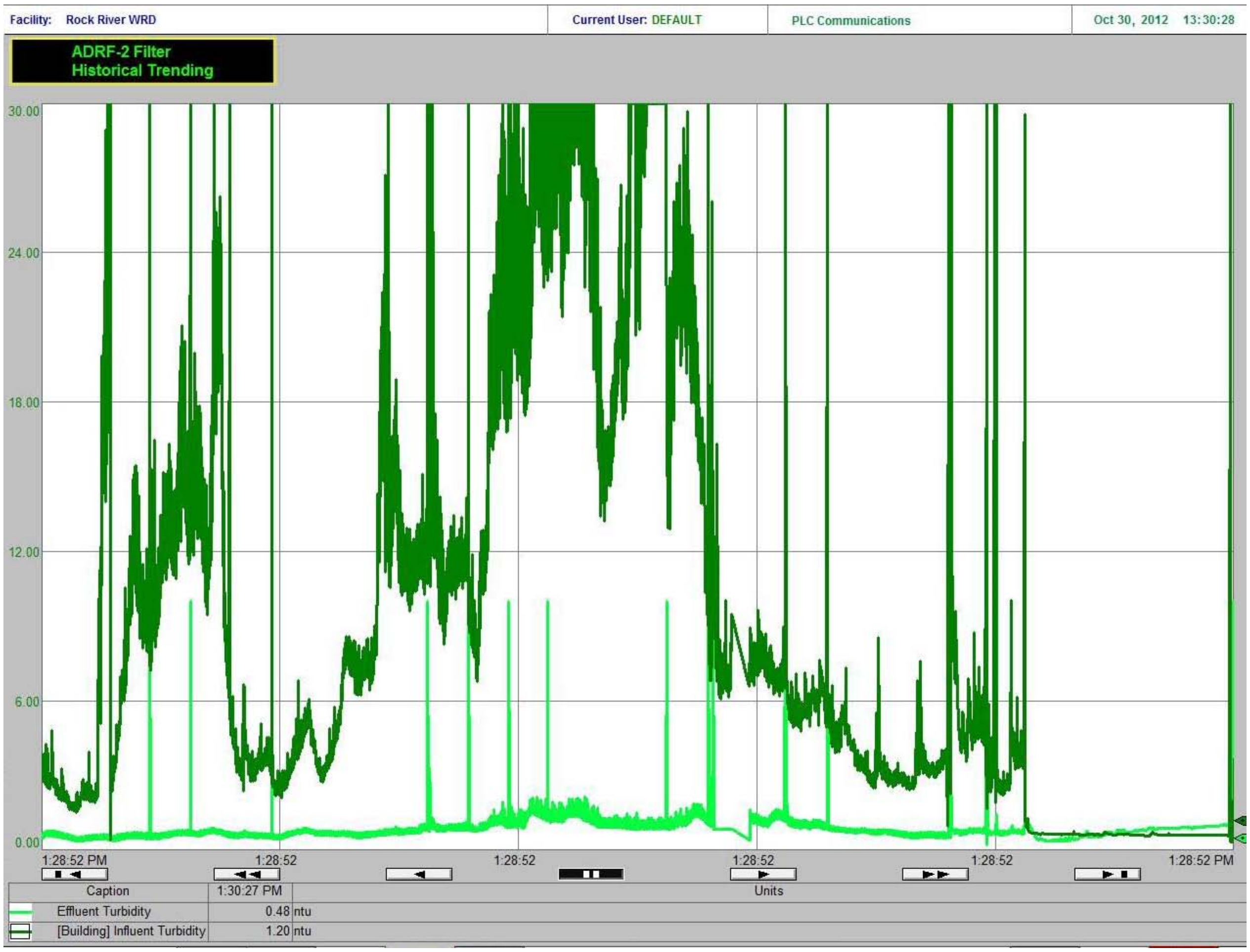

Figure 10. CMF Influent and Effluent Turbidities

Conclusions

Though a UF membrane system can treat secondary settled effluent (SSE) directly and produce a high-quality, low-solids effluent, there are some real advantages to running the SSE through a tertiary filter prior to membrane treatment. The most significant benefit is a 30 – 45% increase in normalized permeability, depending on the membrane flux. The result is a 30 – 45% reduction in the number of membranes used in the system, which may translate into a $130,000 – 205,000 O&M cost savings over the 20-year life of a 1 mgd plant.

Another advantage to including the tertiary filter is its ability to handle large swings in solids loading. Without the filter, the UF system must be able to quickly adjust should a large solids excursion occur with an upstream clarifier or attached growth biological system; these adjustments may include increases in backwash and CEB frequencies and durations, which reduce recovery and increase energy and chemical costs. In addition, more frequent CIP cleanings will be needed for membranes that are exposed to large spikes in influent solids.

Finally, the addition of the tertiary filter gives the system a level of flexibility that doesn't exist when the filter is omitted. If the required final effluent quality can be met without having to run all of the CMF effluent through the UF system, the O&M requirements of the UF system can be reduced accordingly. In contrast, all of the secondary effluent has to be processed through the UF system if the filter isn't included; this will increase the energy and chemical usage and result in shorter membrane life.

References

Koti, Marcel (2011), "Operating Experiences – Tertiary Wastewater", *inge watertechnologies AG – UF Training*, March 8-10, slides 6-8.

Lau, Peter (2008), "Calculation of Flow Rate From Differential Pressure Devices – Orifice Plates", *EMATEM – Sommerschule*, August 2-4, 2008, slide 27.

Zheng, Xing (2011), "Stabilizing the Performance of Ultrafiltration in Filtering Tertiary Effluent – Technical Choices and Economic Comparisons", *Journal of Membrane Science*, volume 366 (2011), pp. 82-91.

The Evaluation of Cloth Media Filtration (CMF) as Pretreatment to UF in Wastewater Reuse Applications

Dave Holland

America's Authority in Membrane Treatment
AMTA
American Membrane Technology Association
Improving America's Waters Through Membrane Treatment and Desalting

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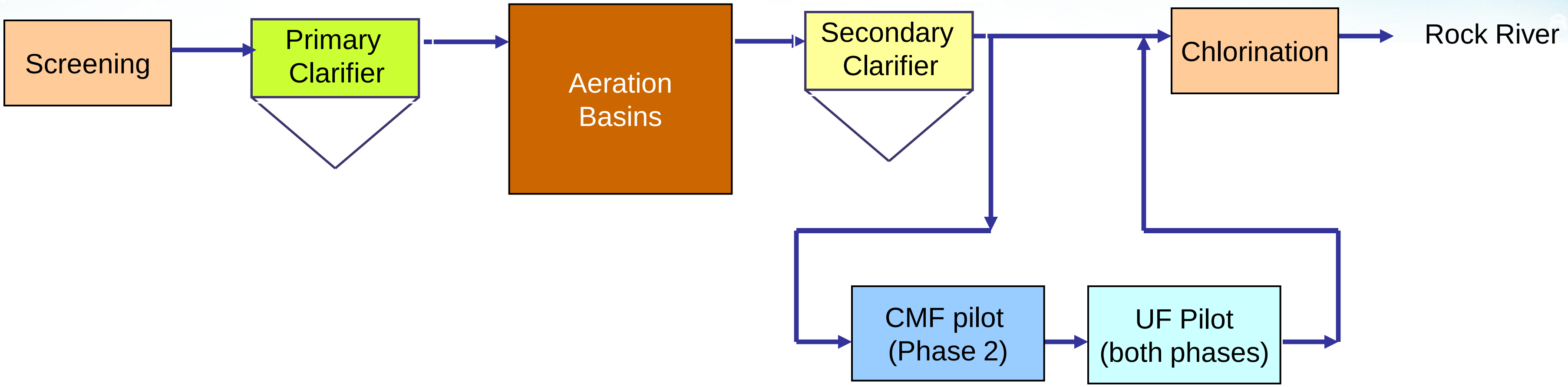
Presentation Outline

- Testing Objectives
- Experimental Facilities
- Key Parameters
- Experimental Design
- Results
- Conclusions

Testing Objectives

- Quantify permeability advantage with CMF
- Compare O&M costs with & without CMF
- Estimate life cycle costs with & without CMF
- Determine pro’s and con’s of adding CMF

Experimental Facilities



- CMF and UF pilot systems treated secondary settled effluent (SSE) at the Rock River Water Reclamation District (RRWRD) in Rockford, IL

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Experimental Facilities

- Research Center at RRWRD
- UF pilot containing single 60 m^2 membrane module with 0.02 µm pores (both phases)
- CMF pilot containing single 5 m^2 filter disk with 5 µm openings (Phase 2 only)

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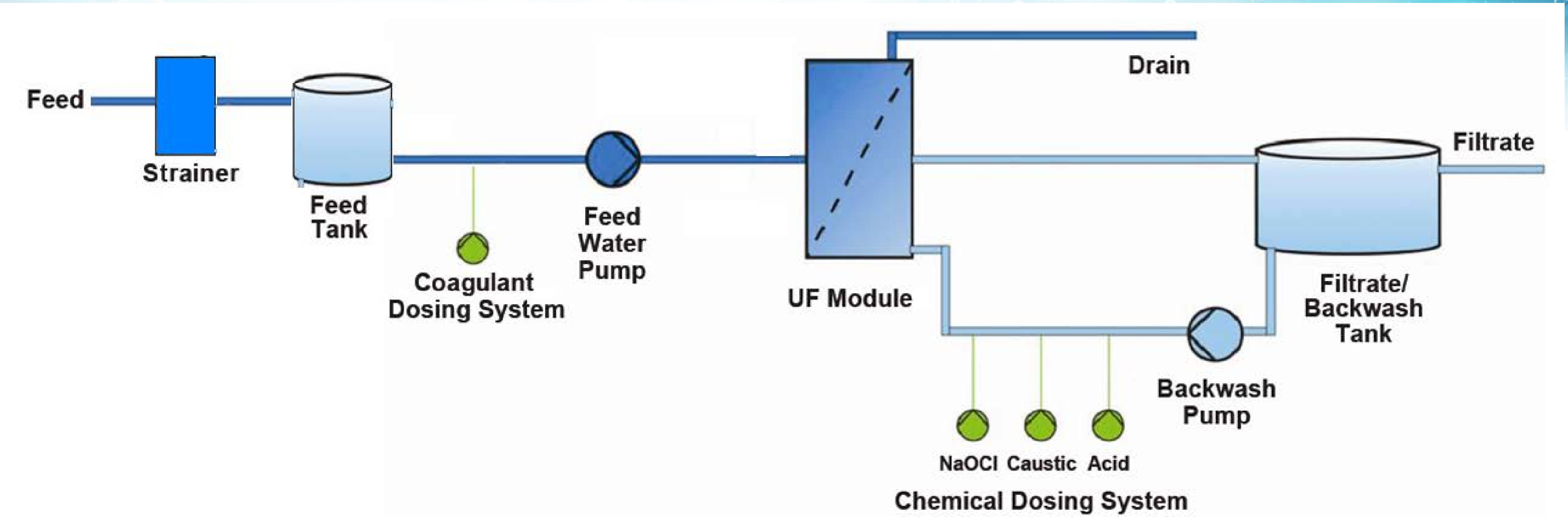
Experimental Facilities – UF Pilot



- Alum injected to combine small particles
- Filtered in dead-end mode (no reject/recycle)
- Cleaned with filtrate and chemicals

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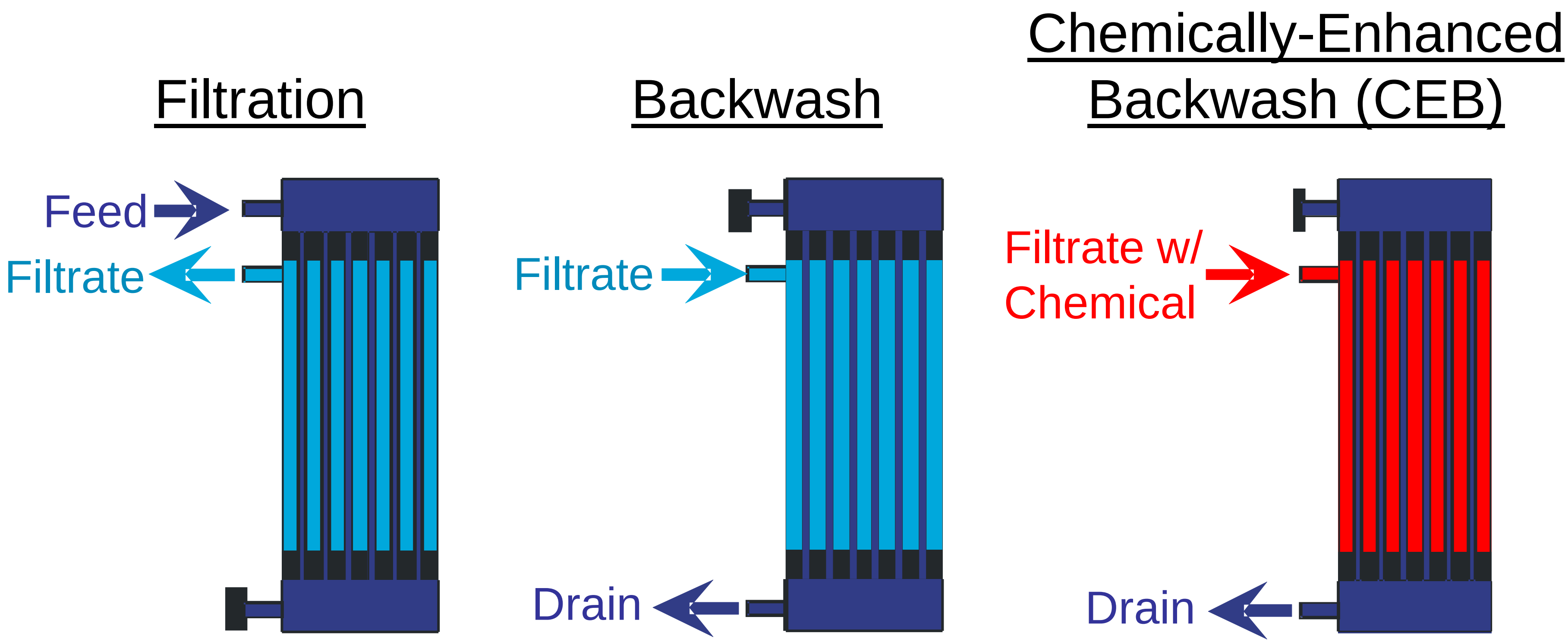
Experimental Facilities – UF Pilot

- (3) Operating Modes



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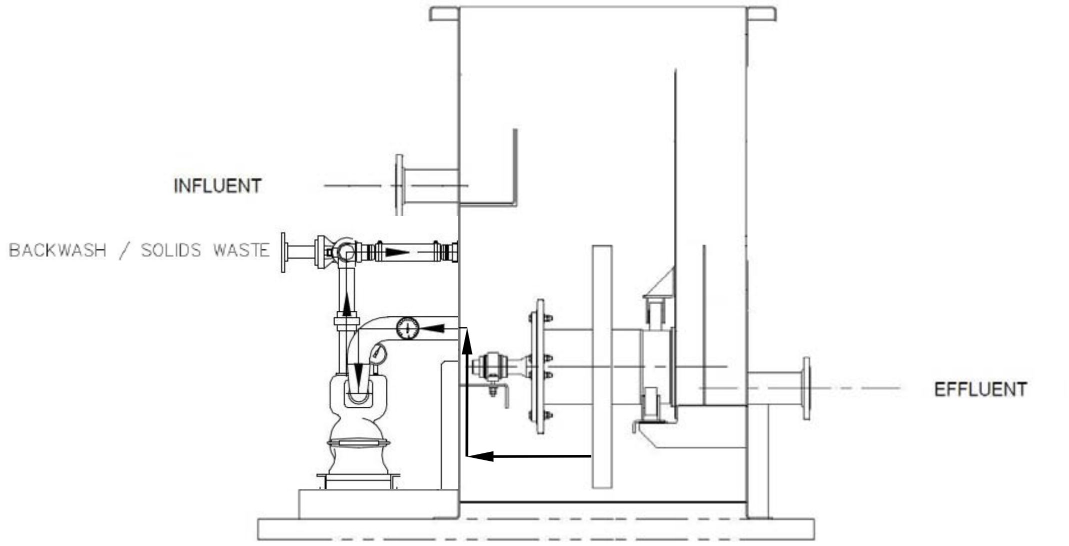
Experimental Facilities – CMF Pilot



- No moving parts during filtration mode
- Effluent weir keeps disk fully submerged
- Filtration continues during backwash

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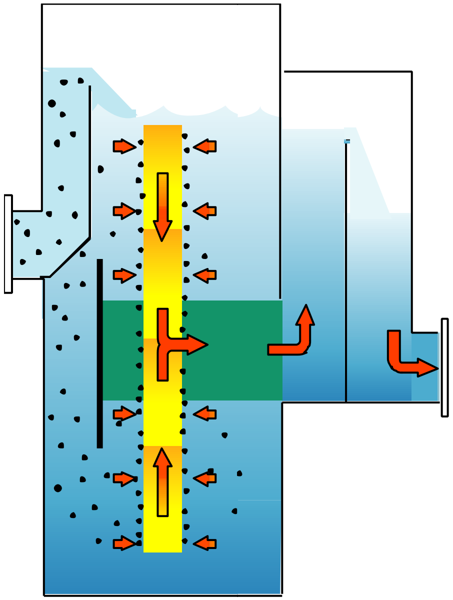
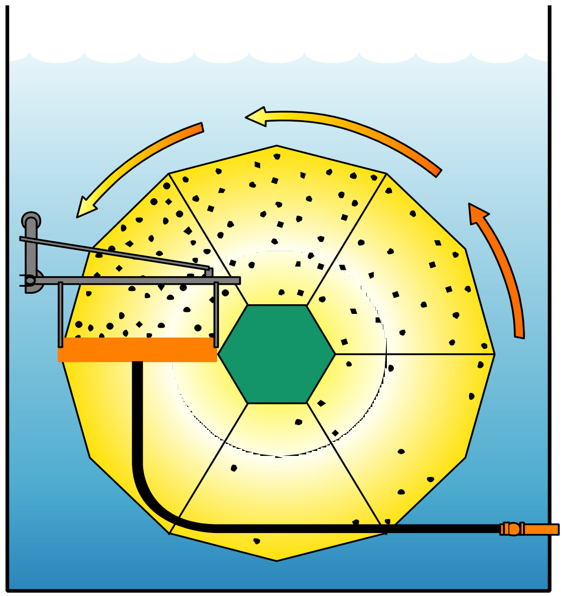
Experimental Facilities – CMF Pilot

- (3) Operating Modes

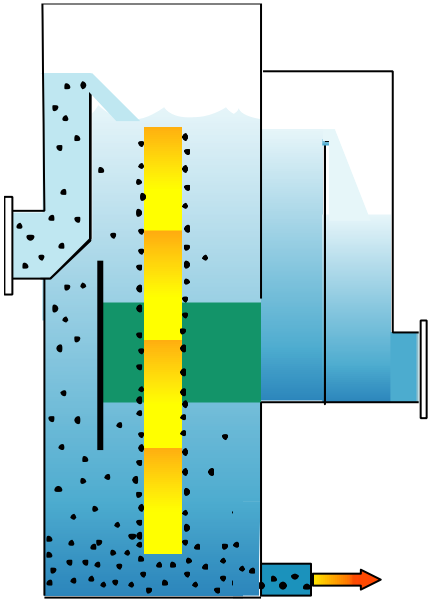
Filtration



Backwash



Sludge Removal



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Key Parameters

- Membrane Flux
- Trans-Membrane Pressure
- Membrane Permeability
- System Recovery

Key Parameter: Membrane Flux (MF)

- The flow Q that passes through a given membrane area A:

$$MF = Q / A$$

- Net flux MF_N uses only the effluent flow Q_E:

$$MF_N = Q_E / A$$

- Gross flux MF_G also takes into account daily filtration time T_F and backwash volume Q_B :

$$MF_G = MF_N \; x \; (1440 / T_F) \; x \; (Q_E + Q_B) / Q_E$$

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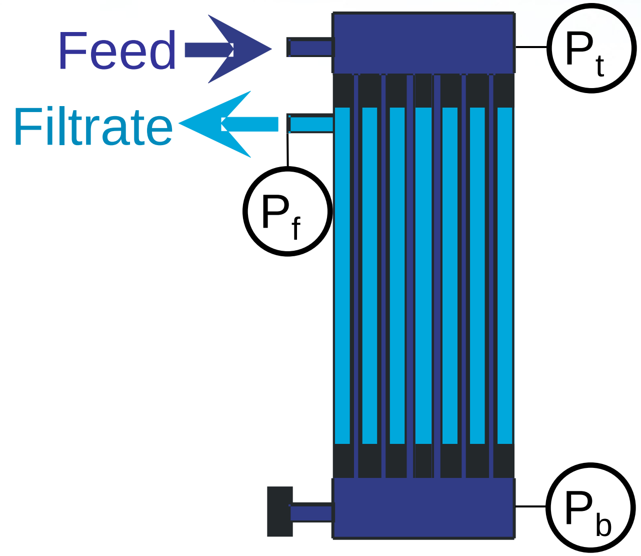
Key Parameter: Trans-Membrane Pressure (TMP)



- The average of top feed pressure P_t and bottom feed pressure P_b less filtrate pressure P_f:

$$TMP = [(P_t + \mathrm{P_b}) / 2] - P_f$$

Key Parameter: Membrane Permeability (MP)

- The gross membrane flux MF_G per unit trans-membrane pressure TMP:

$$MP = MF_G / TMP$$

- **Rule of thumb:**
  - **Doubling the flux will quadruple the TMP and halve the permeability.**

Key Parameter: Membrane Permeability

- Before comparing permeabilities, they should be normalized to the same temperature.
- The normalized membrane permeability MP_n is the membrane permeability at the actual temperature (MP_T) adjusted with temperature correction factor K:

$$MP_n = MP_T / K$$

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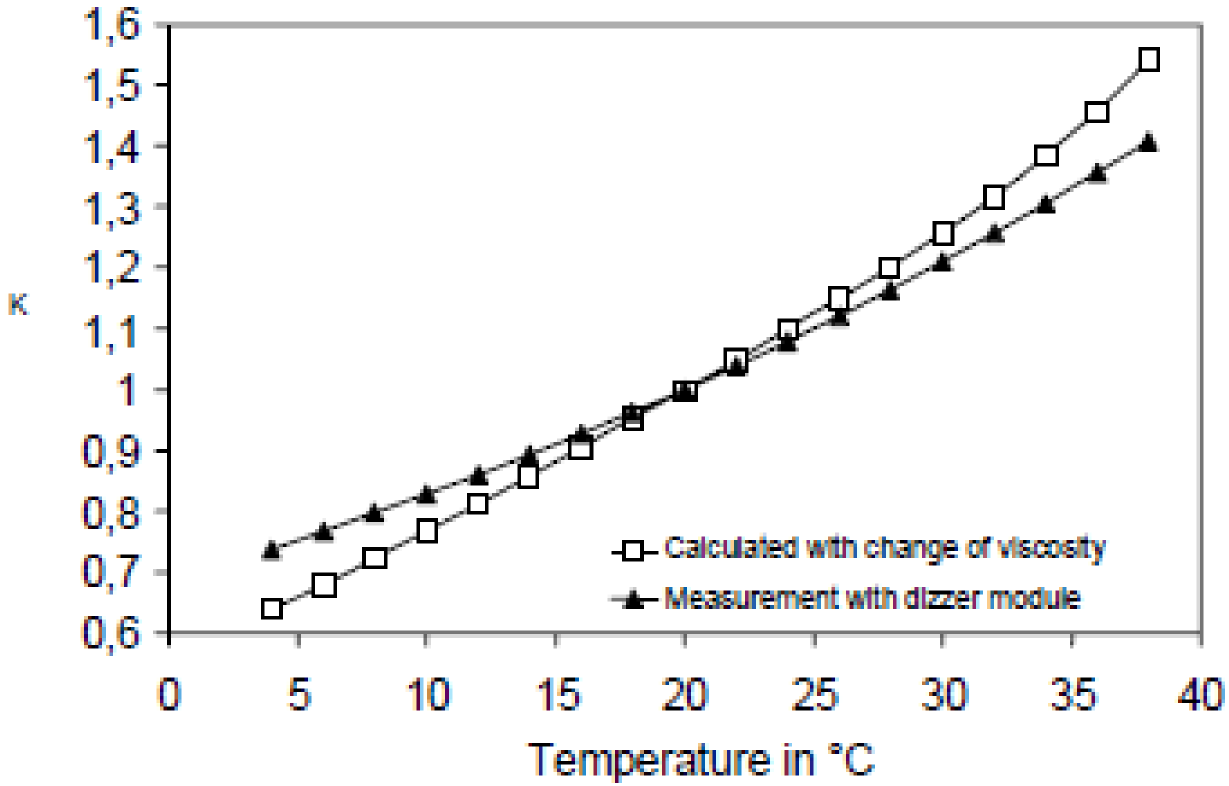
Key Parameter: Membrane Permeability

- Temperature Correction Factor Chart:



Key Parameter: System Recovery (SR)

- The percentage of total influent flow Q_I that actually becomes effluent flow Q_E:

$$SR = (Q_E / Q_I) \times 100\%$$

- Q_I also includes filtrate used for backwashes and chemical dilution.
- A lower SR means more membranes will be needed to get the required Q_E.

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Experimental Design

- Phase 1: Membrane without Filter
  - 5 pilot runs
  - Varied flux, sequence, filtration times, cleaning intervals, and alum dosage/duration
  - Goal was to achieve highest sustainable permeability at highest net flux

Experimental Design: Phase 1

- Filtration Settings

Trial #	Filtration Bottom (min.)	Filtration Top (min.)	Forward Flush Bottom (sec.)	Forward Flush Top (sec.)
1	45	45	30	30
2	30	30	30	30
3	45	45	30	30
4	30	30	30	30
5	30	30	0	0

Experimental Design: Phase 1

- Chemical Settings

Parameter	Units	2/21-3/22	3/22-4/12	4/14-6/26
CEB 1 Interval	Hours	12	8	6
CEB 2 Interval	Hours	999	12	15
CEB 3 Interval	Hours	8	36	48
Alum Dosage	mg/L Al^{+3}	1.5	3.0	5.0
Alum Injection Duration	Type	Continuous	Continuous	Pre-Coat

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Experimental Design

- Phase 2: Filter Followed by Membrane
  - 10 pilot runs
  - Varied fluxes and filtration times
  - Goal was to compare permeabilities with those achieved in Phase 1

Experimental Design: Phase 2

- UF Settings

Trial Number	Gross Flux (lmh)	Feed Flow (gpm)	Filtration Bottom (min.)	Filtration Top (min.)
1	50	13.2	30	30
2	60	15.9	30	30
3	70	18.5	30	30
4	70	18.5	45	45
5	75	19.8	30	30
6	75	19.8	45	45
7	80	21.1	30	30
8	80	21.1	45	45
9	85	22.5	30	30
10	85	22.5	45	45

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Results

- Feed Water Characteristics
- UF Performance During Each Phase
- Comparison of the Two Phases

Results: Feed Water Characteristics

- Phase 1 influent to the UF system:

Parameter	Unit	Min	Max	Average
pH	pH	7.0	7.2	7.1
Temperature	°F	63	69	66
Turbidity	NTU[1]	1.1	2.2	1.3
Total Suspended Solids (TSS)	mg/L	1	7	2.7
Chemical Oxygen Demand (COD)	mg/L	14	38	22.3
5-Day Biological Oxygen Demand (BOD_5)	mg/l	3	9	5.4
Carbonaceous BOD (CBOD)	mg/l	2	7	3.5

[1]Nephelometric Turbidity Unit (NTU) – these values are estimates

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Results: Feed Water Characteristics

- Phase 2 influent to the filter:

Parameter	Unit	Min	Max	Average
pH	pH	6.8	7.0	6.9
Temperature	°F	68	72	70
Turbidity	NTU[1]	1.62	6.38	4.13
Total Suspended Solids (TSS)	mg/L	1	20	8.5
Chemical Oxygen Demand (COD)	mg/L	24	49	34.2
5-Day Biological Oxygen Demand (BOD_5)	mg/l	5	16	9.7
Carbonaceous BOD (CBOD)	mg/l	3	10	5.6

[1]Nephelometric Turbidity Unit (NTU)

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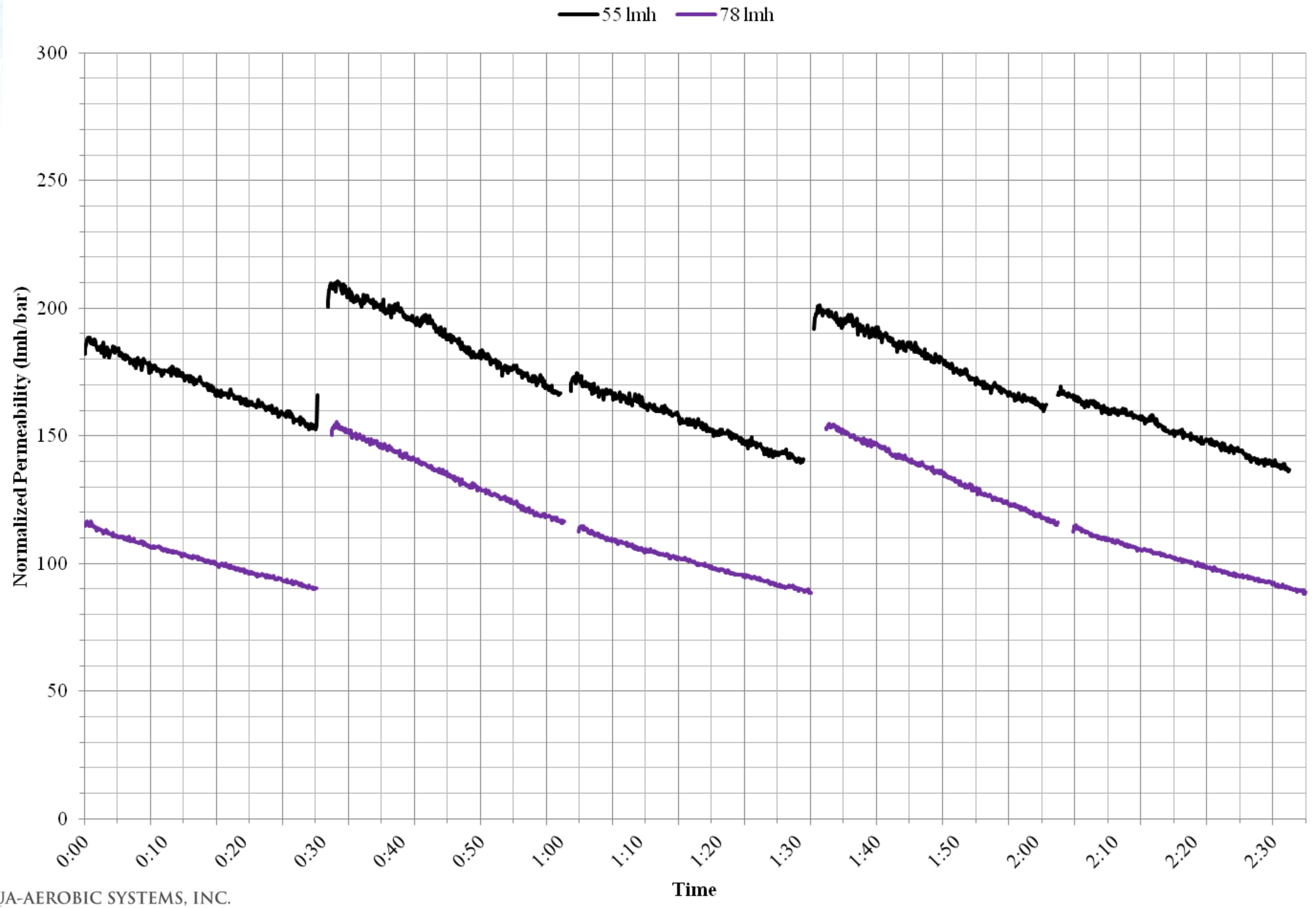
Results: Phase 1 Performance



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Results: Phase 1 Performance

- The average permeability dropped from 170 to 115 lmh/bar (32%) when the flux increased from 55 to 78 lmh (42%).
- At $0.08/kwh, a 32% increase in feed pump power will add about $764/yr/MGD.
- 42% less equipment will result in other O&M savings of about $10,500/yr/MGD[1].
- Estimated annual O&M cost savings is $9,736/MGD.

[1]Based on estimate given by Xing Zheng in *Journal of Membrane Science*, Vol 366, 2011

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Results: Phase 2 Performance

- Filter Effluent:

Parameter	Unit	Min	Max	Average	% Removal
Turbidity	NTU[1]	1.25	2.14	1.69	59
Total Suspended Solids (TSS)	mg/L	ND[2]	8.6	1.3	85
Chemical Oxygen Demand (COD)	mg/L	11	22	18	47

[1]Nephelometric Turbidity Unit (NTU)
[2]Non-Detectable

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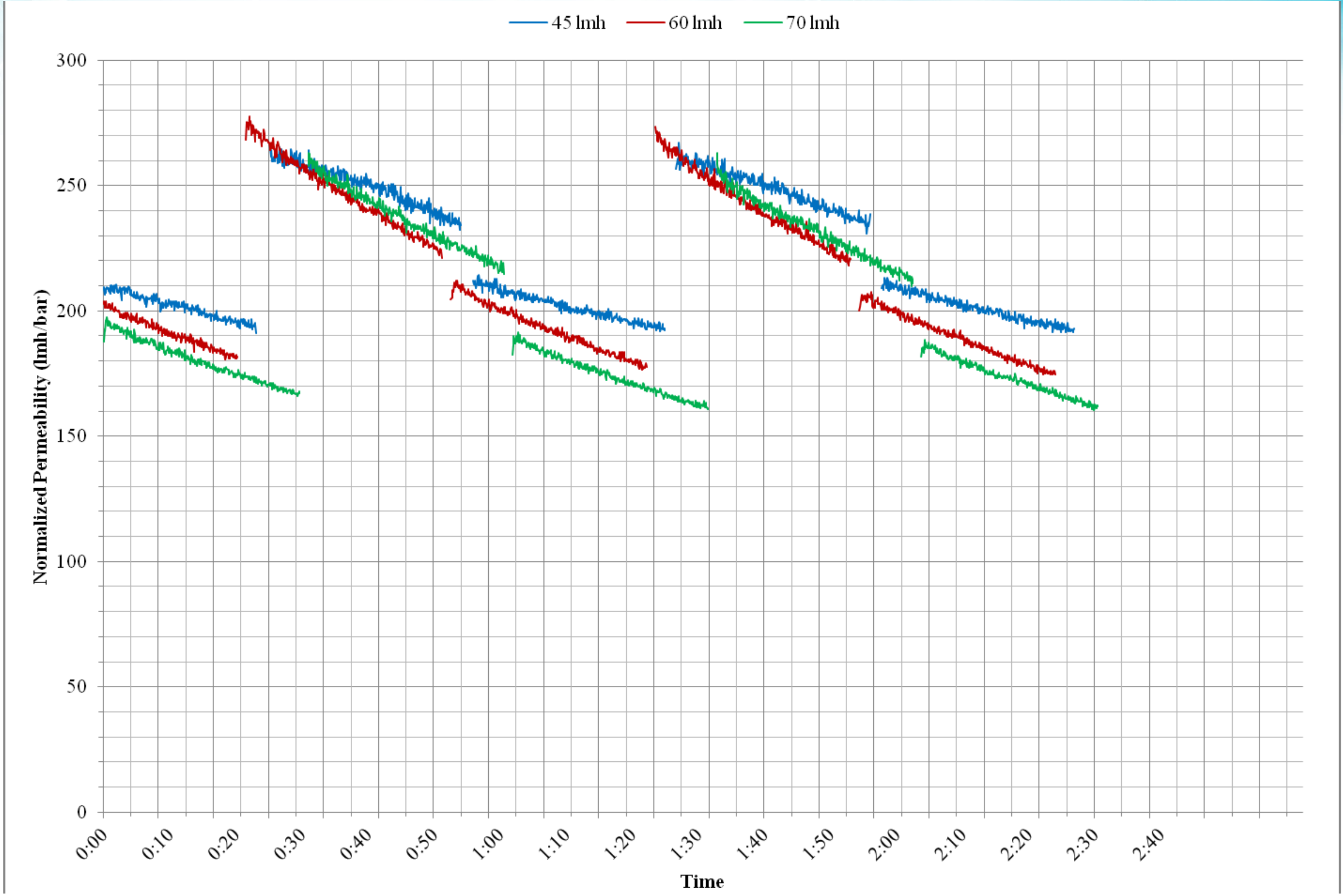
Results: Phase 2 Performance



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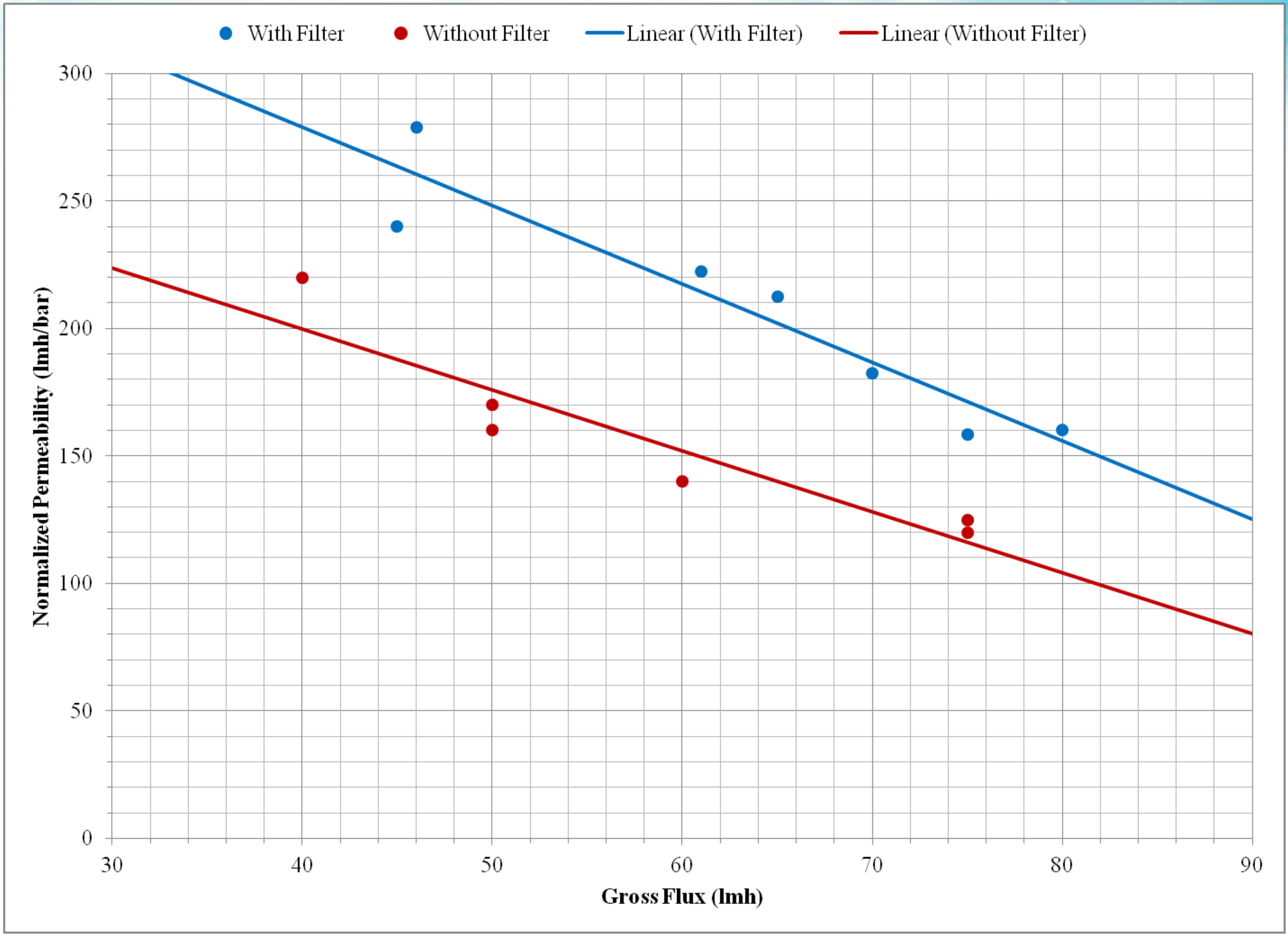
Results: Phase Comparisons

- At 75 lmh gross flux, the 30% higher permeability with the filter results in an estimated annual O&M savings of $6,500/MGD[1].
- At $0.42/gpd[2], the UF system capital cost savings will be about $0.14/gpd, close to the cost of the filter equipment.
- The result is $130,000 savings over 20 yrs.

[1]This is 30% of the $25,000/yr/MGD cost for the UF less $1,000/yr/MGD for the filter.

[2]Based on estimate given by Xing Zheng in *Journal of Membrane Science*, Vol 366, 2011.

Results: Phase Comparisons



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Results: Phase Comparisons

- Two Other Observations:
  - During Phase 2 (with the filter), there were much fewer spikes in TSS to the membranes and clean-in-place (CIP) cleanings were rarely needed.
  - During Phase 2 (with the filter), the turbidity of the filter was frequently lower than 1 NTU; therefore, some plants will be able to bypass the membranes with most or all of the flow and still meet their permit requirements.

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Conclusions

- Running secondary settled effluent through a cloth media filter prior to UF resulted in a 30–45% increase in normalized permeability.
  - About $130-205K less for 20-yr life cycle cost.
- With the filter in front of the UF, the membranes will not see large swings in solids and periodic CIPs will not be needed.
- At times, the required final effluent quality can be met without having to run all of the filter effluent through the UF system.

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For More Information:

- Contact Dave Holland at (815) 639-4470 or dholland@aqua-aerobic.com

Thank You

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