AQUA-AEROBIC SYSTEMS, INC.
A Metawater Company

FULL-SCALE EVALUATION OF CERAMIC MEMBRANE FILTRATION FOR THE CITY OF ASHLAND, OREGON DRINKING WATER TREATMENT PLANT

PILOT TEST REPORT

March 23, 2018

Testing conducted at:

CITY OF ASHLAND, OREGON WATER TREATMENT PLANT (WTP)

1400 Granite St.

Ashland, OR 97520

August 14, 2017 through December 14, 2017

AASI Project #100752A

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Abbreviations

AASI: Aqua-Aerobic Systems Inc.

ACH: Aluminum chlorohydrate

BW: Backwash

CEB: Chemically enhanced backwash

CIP: Clean-in-place

CMF: ceramic membrane filtration

CWFT: Clean water flux test

DOC: Dissolved organic carbon

HMI: Human-machine interface

MFEM: Membrane Filtration Equipment Manufacturer

MWJ: Metawater Japan

PDT: Pressure decay test

PLC: Programmable logic controller

TC: Temperature Corrected

TID: Talent Irrigation District

TMP: Trans-membrane pressure

TOC: Total organic carbon

WTP: Water treatment plant

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1. Objective

Aqua-Aerobic Systems Inc. (AASI) conducted a pilot test to demonstrate the performance of the AASI Ceramic Membrane Filtration (CMF) CM-3 pilot system in treating surface water for potable use at the CITY OF ASHLAND, OREGON WATER TREATMENT PLANT (WTP). This study was conducted from August to December 2017 and included three design runs. The first design run was conducted using water sourced from the Talent Irrigation District (TID). This water is considered to be difficult based on its higher total organic carbon (TOC) and turbidity. The second and third design runs were conducted using water sourced from the Reeder Reservoir. These two runs were distinguished by the second being a conservative run whereas the third run was conducted with an aggressive flux rate. The recorded data will be useful in projecting the full-scale design requirements to ensure sustained and reliable performance. The data from all three design runs are detailed and discussed in this report.

2. Summary

AASI conducted three design run pilot tests at the City of Ashland, OR WTP using the CM-3 ceramic membrane pilot system. This report presents the results from all three design runs conducted at the Ashland WTP. Commissioning began during the week of August 14, 2017, and the first optimization period was conducted during the week of August 21, 2017.

Design run 1 began on August 30, 2017 and was completed on September 27, 2017. Operating conditions established during the optimization period for design run 1 were set at a flux of 50 gallons per ft^2 membrane area per day (gfd), 45 minute filtration/backwash (BW) cycle times, and daily chemically-enhanced backwashes (CEBs) with one-hour soak. The starting baseline trans-membrane pressure (TMP) value was 3.5 psi, and ended at a lower TMP value of 2.2 psi (most likely attributed to the 2 ppm coagulant added to the pilot influent toward the end of the run, which was the only coagulant being added during run 1). TMP peaks (i.e., max TMP value between backwash events) ranged from 10-25 psi.

A pressure decay test (PDT) was conducted daily; despite having some program issues in the first two weeks, the PDT was functional in the second half of design run 1. Total system recovery was 97%. Total reject rate (i.e., BW and CEB waste volumes as a percentage of feed volume) was 3%. Feed turbidity ranged between 5 and 20[1] NTU. Filtrate turbidity ranged between 0.01 and 0.09[1] NTU, and averaged 0.018 NTU. Feed pH remained stable at 7, and feed temperature ranged from 50 to 65 degrees Fahrenheit. Lastly, a clean-in-place (CIP) cleaning was performed prior to starting the 28 day test and directly after the test.

Design run 2 was conducted from October 3 to October 31, 2017. Operating conditions established during the optimization period for design run 2 were set at a flux of 50 gfd, 60-minute filtration/BW cycle time (later changed to 720 min), and daily CEB with one-hour soak. Starting baseline TMP value was 1.8 psi, with peaks ranging from 2–12 psi. Filtration time (i.e., BW frequency) was set at 60-minute intervals from 10/3 to 10/23/2017, but was later changed to 720-minute (12-hour) intervals from 10/23 to 10/31/17. The reasoning for this change was to increase percent recovery. Total system recovery was 98.3%. Total reject rate was 1.7%. Feed turbidity ranged between 0.2 and 2[1] NTU. Filtrate turbidity ranged between 0.002 and 0.05[1] NTU, and averaged 0.013 NTU. Feed pH averaged 7.3, and feed temperature ranged from 49–57 degrees Fahrenheit. A CIP was conducted upon completion of this test run.

Design run 3 was conducted from November 14 to December 12, 2017. Operating conditions established during the optimization period for design run 3 were set at a flux of 150 gfd, 90-minute filtration/BW cycle time, and a CEB every 3 days with a one-hour soak. Starting TMP value was 5.4 psi, with peaks ranging from 9.4–17 psi. Filtration time was set at 90-minute intervals. Total system recovery was 99.5%. Total reject rate was 0.5%. Feed turbidity ranged between 0.6 and 3[1] NTU. Filtrate turbidity ranged between 0.0018 and 0.08[1] NTU, and averaged 0.010 NTU. Feed pH averaged 7.1, and feed temperature ranged from 38–48 degrees Fahrenheit.

[1]The turbidity values are filtered to exclude the spiked values which were later determined to be due to air compressor induced vibrations. The filtrate turbidimeter was removed from the skid on 9/15/17. Spiking events were nearly eliminated; the few remaining spiking events were attributed to bubbling from vibrations affecting the filtrate tank. A bubble trap was installed on the feed side of the filtrate turbidimeter unit to minimize bubbling effects. This was done prior to starting design run 3.

During the last part of run 2 (after 10/11/17) and all of run 3, the Owner was injecting 8 ppm of 50% aluminum chlorohydrate (ACH) coagulant into the raw water upstream of the three MFEM pilot systems. Coagulation enhances ceramic membrane filtration by reducing fouling potential and facilitates operation at a higher flux. In addition, all three runs operated with daily pressure decay tests (PDTs) (refer to Appendix C). Feed and filtrate turbidity data, feed pH, and feed temperature data were collected as well. Operating conditions and performance data from all test runs are detailed in section 5.

An average of 4.28 LRV was achieved during run 3, with a max LRV of 4.73 and a min LRV of 4.00. We encountered many difficulties with the PDT during run 1, both mechanically (i.e., air leaks via piping) as well as programmatically. Eventually, both issues were resolved for part of run 1. However, PDT data collection did not occur until Run 3 due to a program set up issue.

A clean-in-place (CIP) membrane cleaning was performed before and after each design run to demonstrate recovery of permeability, the results of which are presented in Section 5; refer to Appendix A for the CIP procedure used. The CIP consisted of two phases: a) 1% Citric Acid recirculation for 2 hours, followed by a 2 hour soak; and b) 0.03% NaOCl recirculation for 2 hours, followed by a 2 hour soak. The use of these chemicals, and their corresponding concentrations and membrane-exposure durations were established by Metawater as to be the most effective and efficient in restoring membrane baseline / permeability condition.

The chemical solutions for the first and second CIPs were heated to 80 deg Fahrenheit, but the third CIP was not heated.

To mitigate bio-fouling and enhance performance, a chemical-enhanced backwash (CEB) with one-hour soak was conducted daily during design run 1, weekly during part of design run 2, and every 3 days during design run 3; refer to Appendix B for the CEB procedure used. The CEBs alternated between two different sequences: one consisted of a 20-minute soak in hydrochloric acid (HCl, alternatively, H2SO4 can be used) at pH 1-2 followed by a 40-minute soak in sodium hypochlorite (NaOCl) at 0.01-0.03% (100–300 ppm), and the other consisted of a 60-minute soak in 0.01-0.03% NaOCl. Operational data were logged daily and can be provided upon request. A summary of the operating conditions are shown in Table 3-1. CEB

chemicals were not heated at any time throughout testing. The CEB recipe stated above has been established by Metawater on similar water applications. The CEB recipe can be optimized further for higher performance and/or lower consumption.

Based on these results, we recommend the full-scale system consist of (2) 12-row units with (10) modules per row, which will provide a total of 6,000 m^2 (64,590 ft^2) of membrane area, resulting in a flux of 124 gfd with all rows in service at 8 MGD and a flux of 129 gfd with one row out of service (each row has its own set of valves such that it can be operated totally independently of the other rows). We feel this approach is best for the following reasons:

- During Run 3, the pilot maintained stable operation at 150 gfd even though the Reeder Reservoir water TOC was the same as that of the TID source (2.9 ppm) and the water temperature was as low as 38 degrees Fahrenheit at a time of year when flows are typically less than 30% of design[1]. It's expected that even higher fluxes are possible in the summer months at higher water temperatures.
- Even in the drought of 2014, the TID water source made up only 17% of the total supply[2], most of it used during the warm summer months when higher fluxes are possible.
- During Run 1, the pilot maintained stable operation at 50 gfd even though the TID water turbidity was as high as 20 NTU at 54 degrees Fahrenheit and there was little to no coagulant addition. This run also demonstrated that operation became much more stable with as little as 2 ppm of 50% ACH (0.25 ppm as Al); we would expect to run at a much higher flux had the coagulant dosage been at the 8 ppm (as 50% ACH) that was used during Run 3 and the end of Run 2.

Based on the conditions used in Run 3 (at the lowest water temperatures), we recommend using the following operating parameters during the 7 colder months. Note that the backwash/filtration and CEB intervals are double those

[1] From Figure 4.3 of the City of Ashland Comprehensive Water Master Plan, adopted April 17, 2012.
[2] See the data for 2014 on Figure 1 in Appendix A of the WATER QUALITY ANALYSIS AND TREATMENT PROCESS SELECTION, drafted July 2017

used during Run 3 and the CIP interval is double that predicted during the run; the reason for this is that the flow during the winter months is typically about half of the design flow[3]. We feel this is still conservative since the actual membrane area provided is 19% higher than required by the Run 3 flux at design flow.

- Backwash/filtration cycle interval = 215 minutes
- CEB interval = every two days
- CEB soak time = 60 minutes (same used in Run 3)
- CEB sequence = alternate between 1-2 pH HCl-0.03% NaOCl and 0.03% NaOCl only. These are the same CEB parameters used during Run 3, which significantly reduced the TMP.
- CIP interval = when TMP baseline reaches 15 psi, expected to be every 4 months. This is based on the TMP increase during Run 3 and the lower winter flow. If possible, conduct CIP during a lower water demand period for less operator maintenance.
- CIP sequence = 4-hour exposure (2 hours recirculation and 2 hours soak) to each chemical
- CIP chemicals = 1% citric acid followed by 0.3% NaOCl

Based on the conditions used in Run 3 and a 22% decrease in TMP due to the water viscosity difference between the average water temperatures during Run 3 (6 degrees C) and in the summer months (15 degrees C)[4], we recommend using the following operating parameters during the 5 summer months. Again, we feel this is still conservative since the actual membrane area provided is 19% higher than required by the Run 3 flux at design flow.

- Backwash/filtration cycle interval = 110 minutes
- CEB interval = daily
- CEB soak time = 60 minutes (same used in Run 3)

[3] From Figure 4.3 of the City of Ashland Comprehensive Water Master Plan, adopted April 17, 2012.
[4] From Figure 8 in Appendix A of the WATER QUALITY ANALYSIS AND TREATMENT PROCESS SELECTION, drafted July 2017.

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- CEB sequence = alternate between 1-2 pH HCl-0.03% NaOCl and 0.03% NaOCl only. These are the same CEB parameters used during Run 3, which significantly reduced the TMP.
- CIP interval = when TMP baseline reaches 15 psi, expected to be every 90 days. This is based on the TMP increase during Run 3 and the lower baseline TMP expected in the summer months. If possible, conduct CIP during a lower water demand period for less operator maintenance.
- CIP sequence = 4-hour exposure (2 hours recirculation and 2 hours soak) to each chemical
- CIP chemicals = 1% citric acid followed by 0.3% NaOCl

3. Background

The pilot test occurred on-site at the City of Ashland, OR water treatment plant (WTP). The purpose of pilot testing was to evaluate the performance of the AASI CM-3 ceramic membrane filtration system which would be integrated into the implementation of a new WTP for The City of Ashland, OR.

The feed supply to the membrane filtration systems was sourced from both the Reeder Reservoir and Talent Irrigation District. The pilot tests were conducted outdoors under a tent at The City of Ashland, OR, WTP, at 1400 Granite St, Ashland, OR. A simplified diagram of the test site layout is shown in Figure 3-1.

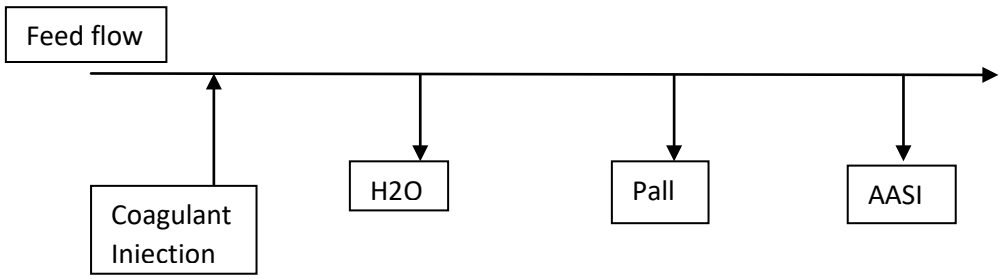


Figure 3-1: Simplified Diagram of Pilot Test Site Layout

Table 3-1: Summary of Pilot Test Operating Conditions

Parameter	Design run 1 (8/30/2017 to 9/27/2017)	Design run 2 (10/03/2017 to 10/31/2017)	Design run 3 (11/14/2017 to 12/12/2017)
a) Coagulant injection	2 ppm 50% ACH into CM-3 unit mix tank[5] From 9/16 to 9/27/17. Prior to this and during, no other coagulant was dosed	8 ppm 50% ACH(Owner-injected) from 10/ 12[5] to 10/31; No coagulant was added from 10/03 to 10/12	8 ppm ACH (Owner-injected) – added throughout test run as common to all MFEMs
b) Filtration cycle /Backwash interval	45 minutes	60 minutes (increased to 720 minutes for the last 1/3 of test)	90 minutes
c) Flux (gfd)	50	50	150
d) Feed flow (gpm)	9.3	9.3	28
e) CEB cycle frequency	1x/day, alternating between acid-oxidant and oxidant only	1x/day Oct 3-22, 1x/week Oct 23-31 (all alternating)	1x/3 days, alternating between acid-oxidant and oxidant only
f) CEB soak time	60 minutes total (split between acid and hypo CEBs)		
g) Acid CEB	HCl to 1-2 pH, 20-minute soak every other CEB; not heated		
h) Oxidant CEB	0.03% NaOCl , alternating between 40- and 60-minute soak; not heated		
i) CIP cycle	28 days		
j) Acid CIP	1% citric acid for 4 hours (2-hour recirculation, 2-hour soak) heated to 80° F after test runs 1 and 2, but not after test run 3		
k) Oxidant CIP	0.3% NaOCl for 4 hours (2-hour recirculation, 2-hour soak), heated to 80° F after test runs 1 and 2, but not after test run 3		

[5] The intent of the WTP was to inject 8 ppm of 50% ACH upstream of the three MFEM pilots, but it was found on October 12 that no coagulant had been injected due to malfunction of the plant's ACH feed system.

3.1 Raw water quality

3.1.1 Run 1 water quality

Feed turbidity ranged between 5 and 20[6] NTU. Filtrate turbidity ranged between 0.01 and 0.09[6] NTU, and averaged 0.018 NTU. Feed pH remained stable at 7, and feed temperature ranged from 50–65 degrees Fahrenheit. Feed TOC ranged from 2.4 to 2.9, feed UV-254 transmittance ranged from 0.054 to 0.077 cm^{-1}, and the feed algae count on September 6 was 190 per ml; filtrate levels were not measured for these parameters.

3.1.2 Run 2 water quality

Feed turbidity ranged between 0.2 and 2[7] NTU. Filtrate turbidity ranged between 0.002 and 0.05[7] NTU, and averaged 0.013 NTU. Feed pH averaged 7.3, and feed temperature ranged from 49–57 degrees Fahrenheit. Feed TOC ranged from 1.4 to 1.9, and feed UV-254 transmittance ranged from 0.027 to 0.043 cm^{-1}; filtrate levels were not measured for these parameters.

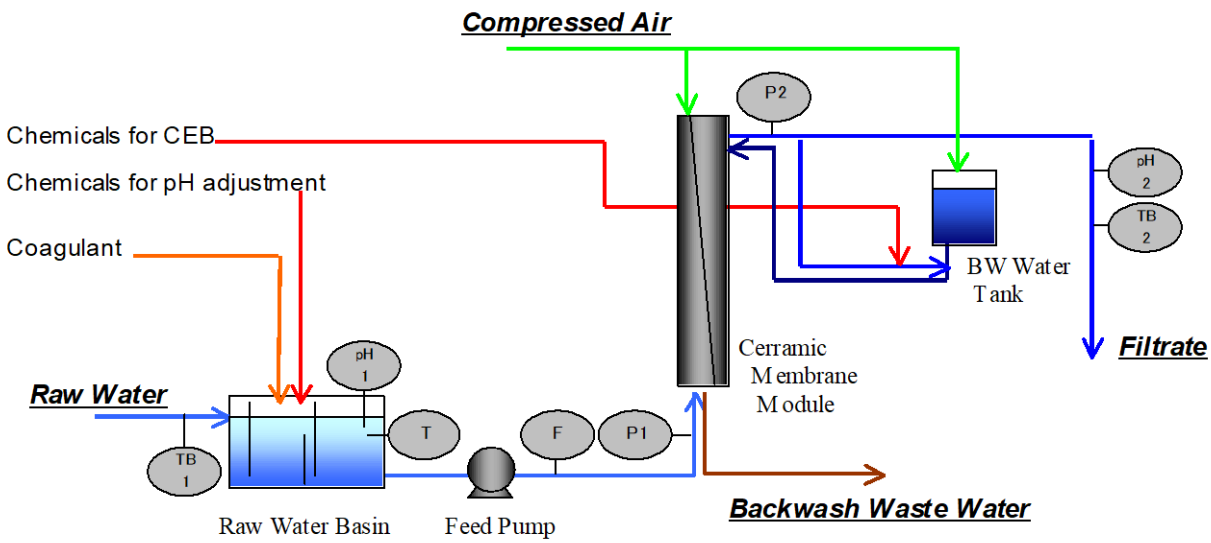
3.1.3 Run 3 water quality

Feed turbidity ranged between 0.6 and 3[8] NTU. Filtrate turbidity ranged between 0.0018 and 0.08[8] NTU, and averaged 0.010 NTU. Feed pH averaged 7.1, and feed temperature ranged from 38–48 degrees Fahrenheit. Feed TOC ranged from 2.2 to 2.9, and feed UV-254 transmittance ranged from 0.047 to 0.087 cm^{-1}; filtrate levels were not measured for these parameters.

[6] See footnote on page 7.
[7] See footnote on page 7.
[8] See footnote on page 7.

4. Test Equipment and Material

Figure 4-1 shows a schematic diagram of the CM-3 pilot system. The pilot system is equipped with a ceramic membrane that was used for the pilot test.



*A strainer was not used on the CM-3 system during testing.

Figure 4-1 Schematic Diagram of the CM-3 Pilot Testing System

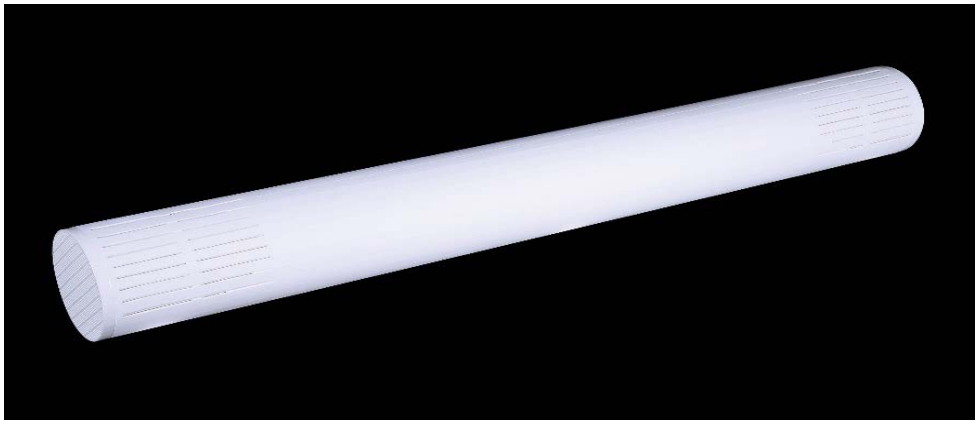
The AASI/MWJ ceramic membrane element is made of alumina-based ceramic material. Its robust properties allow it to operate with a longer membrane life than other membrane types due to its high chemical and thermal resistance and mechanical strength. Higher concentrations of chemical are applicable for cleaning which permits consistent permeability recovery over extended operational periods. Due to a tight pore size distribution, the AASI/MWJ ceramic membranes have higher permeability and less pressure loss across each membrane unit in comparison to polymeric membrane media.

The specifications of the ceramic membrane element used in the study are presented in Table 4-1, and are further illustrated in Picture 4-1. The monolith body of each ceramic membrane measures about 7.1 inches diameter and 59.1 inches long. The nominal pore size is 0.1 micron with a total of 269 ft^2 membrane

surface area. The ceramic membrane operates by inner pressure and dead-end mode filtration. Feed water is supplied to the membrane filtration feed channel and filtered by a separation layer formed on the inside wall of the channel. Filtered water penetrates through the support layer and flows to the filtrate side through the water collection slits. The ceramic membrane has two distinct features when compared to polymeric membranes: (1) a high mechanical strength body that allows high-pressure backwash of 72.5 psi, (2) a large flow channel with diameter of 0.1 inch that can retain particles on the membrane surface during filtration and be flushed out with high-pressure backwash.

Table 4-1: Specifications of Ceramic Membrane

Parameter	Units	Value
Size of Element	mm	180 x 1500
Inside Diameter of Channels	mm	2.5
Active Membrane Area per Module	m^2	25
Flow Direction		Inside out
Number of Channels per Module		2,000
Available Operating Modes		Dead End
Membrane Material		Modified Al_2O_3
Nominal Membrane Pore Size	Micron	0.1
Acceptable Range of Operating pH	S.U.	1-12



Picture 4-1: Ceramic Membrane

A variety of pretreatment options are available to enhance membrane performance and water quality. Applicable pretreatment includes: coagulant (e.g., alum, polyaluminum chloride [PACl], ACH, Ferric Chloride, polymer, etc.), ozone, pre-chlorine, Powdered Activated Carbon (PAC), pH control, etc. Although PACl and ACH have been thoroughly evaluated and established by Metawater to be the most effective coagulants for ceramic membrane in these applications, other coagulant types could have been tested if given more time.

Each ceramic membrane element is installed in a dedicated, stainless steel housing collectively identified as a membrane module. The ceramic membrane module is compliant with NSF/ANSI 61 health effects criteria for drinking water system components. Every single membrane is integrity tested (by bubble point) before shipping, conducted by AASI/MWJ/NGK Quality Assurance/Quality Control (QA/QC). The following outlines the MWJ QA/QC process:

AASI/MWJ/NGK QA/QC process:

Check points:

- Dimensions are measured for every single membrane element.
- Specific flux is measured for every membrane and is calculated by flow rate and trans-membrane pressure.

- Membrane integrity testing is performed via Bubble Testing (visual observation of bubble formation). Each membrane is pressurized at 135 kPa (19.6 psi) and inspected for absence of air leakage.
- Visual inspection is conducted on every single module. Items include color, seal defect, crack, and chipping.

Module and rack (at membrane module fabricator site)

Defined:

- Module: ceramic membrane and housing
- Rack: Module and manifold piping

Check points:

- Material confirmation to mill sheet.
- Visual inspection of items includes standardized evaluation of welding and surface quality.
- Each module and rack assembly is measured for conformance to AASI / MWR quality control specifications.
- Pressurized testing is performed on the rack system to confirm the integrity of air- water surfaces. Each rack system is pressurized with water including module housing and manifold piping. The membrane is not installed at the time of pressurized testing. The system is pressurized up to 750 kPa (108 psi) for 30 minutes and no water leak allowed.
- Air tight testing is conducted on each module to confirm integrity of seal material. Module is hydraulically pressurized to 120 kPa (17.4 psi) from filtrate side in the same way as integrity testing and visually inspected for air leak from seal.
- Completed rack assembly is then leak tested using pressurized by water up to 500 kPa (72 psi).

QA/QC for Membrane filtration skid (at skid fabrication site)

Defined:

- Membrane filtration skid includes: valve unit, interconnection piping.

Check point:

- Air tight testing is conducted on each skid to confirm integrity of seal material all connections. Skid is pressurized up to 300 kPa (44psi) with air. Leak monitoring and visual inspections are conducted as well.

Reporting

AASI / MWJ generates an inspection report identifying the following:

- Rack dimensions
- Membrane specification (visual inspection, dimension, permeability [i.e., clean water specific flux])
- Pressurized and air-tightness testing
- Mill sheet (Rack, skid interconnection piping)

4.1 Operation and Maintenance

When installed in their stainless steel housings, ceramic membranes are extremely durable, difficult to damage, and can maintain stable operation despite fluctuations in water quality. As a result, operation and maintenance of the ceramic membrane system is quite simple and generally easy. For example, unmanned operation has been achieved in many municipal water treatment plants, including WADASHIMA WTP (2.64 MGD) in Japan because of proven reliability over extended operational periods.

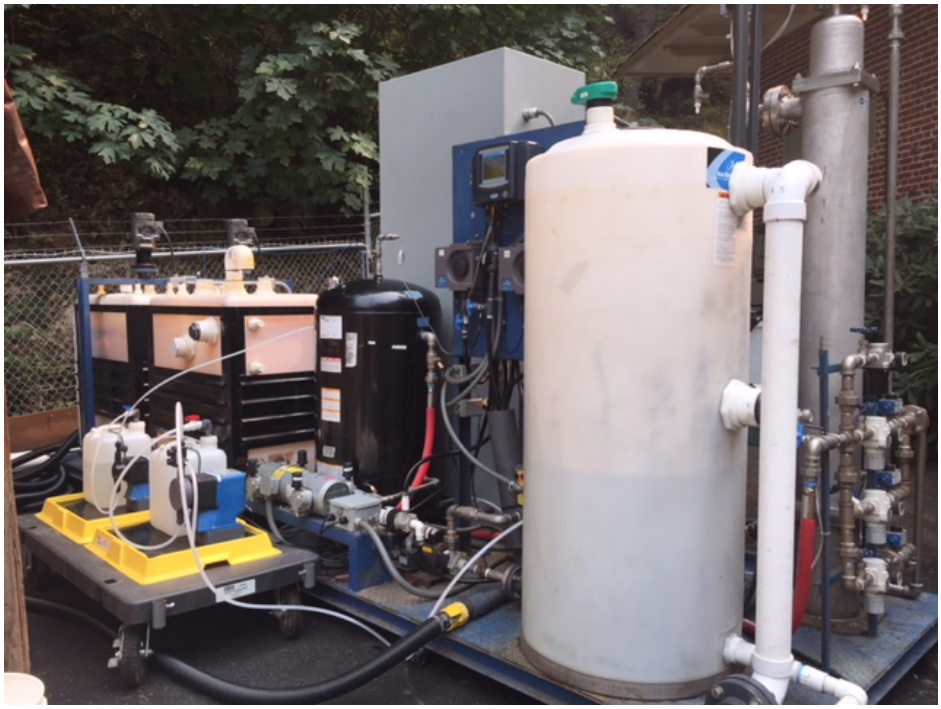
The AASI/MWJ ceramic membrane CM-3 pilot system includes the following components:

- Feed pump
- Pre-treatment process tank (coagulation and flocculation)
- Ceramic membrane module
- Filtrate storage tank
- Backwash water tank
- Backwash waste neutralization tank
- Chemical addition systems for both sodium hypochlorite and acid
- Automated operation: filtration, backwash, CEB, PDT
- Touch-screen user interface
- Air compressor for backwash and pneumatic valve control
- Data logging and remote access capability

The main pilot system is skid-mounted and is equipped with a commercialized ceramic membrane module, as shown in Pictures 4-2 and 4-3.

The AASI / MWR ceramic membrane CM-3 pilot system is a compact, single module unit which is fully functional and representative of a full scale, commercial installation. It is automated for backwashing, chemical enhanced backwashing (CEB), and pressure decay testing (PDT). System operation is controlled by an Allen-Bradley PLC and is interfaced with a HMI (human-machine interface) and remote access capability. All processes and valve operation can be controlled manually and or automatically (with the exception of the CIP process, which is conducted manually). The complete pilot system consists of the main membrane skid, a feed pump and mix tank skid, and a chemical skid.

Picture 4-2: AASI CM-3 Pilot Unit – Front view



Picture 4-3: AASI CM-3 Pilot Unit – Rear View

4.2 Process Description

The CM-3 pilot system has two operational sequences: filtration and backwash. Raw water flows into a receiving part of the coagulation tank consisting of two mixing tanks in series: rapid mixing and slow mixing tank. A measured amount of coagulant is dosed at the rapid mixing tank injection port. Total detention time at 18.6 gpm is 12 minutes. During the filtration process, feed water is pumped to the bottom of the membrane module and enters the membrane channels through inside to out flow path as shown in Figure 4-2 (filtration stage).

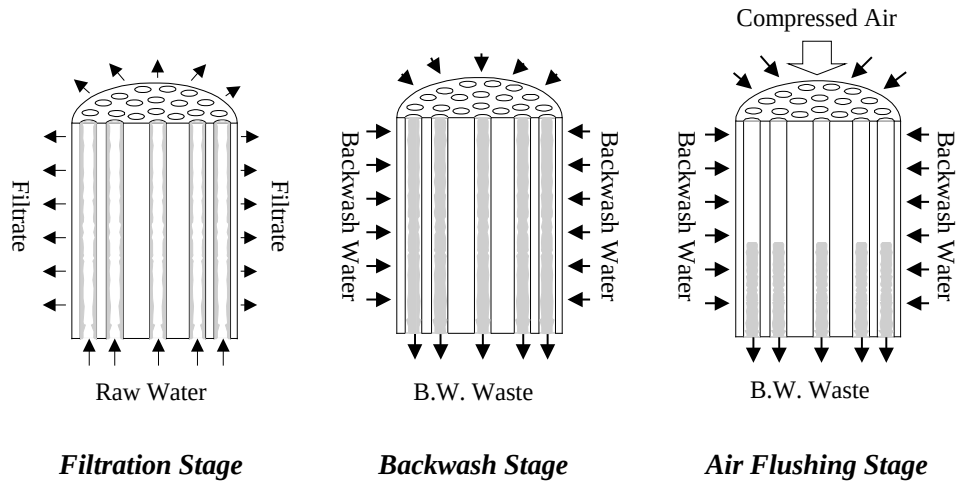


Figure 4-2: Filtration and Backwash

At completion of the filtration process, the backwash (BW) process is initiated to recover membrane permeability. The backwash tank is filled to a prescribed volume with filtrate prior to the backwash process. Prior to backwash initiation, approximately 72.5 psi of air pressure is applied to the backwash tank. This high pressure creates reverse water flow which dislodges accumulated solids from the membrane surface. An immediate air flush step discharges the solid material outside of the membrane channels. Reverse flow and air flush operations occur within 10-20 seconds. Total down-time associated with the backwash process is about 1-3 minutes and permits a very high membrane recovery. The filtration, backwash and air-flushing processes are shown in Figure 4-2. The backwash process can prevent blockage of the membrane to ensure stable operation over an

extended period of time. During the filtration process, a cake layer is formed on the membrane surface that results in reduction of specific flux (pressure normalized flux). After backwashing, the specific flux is recovered to the initial specific flux level.

For general maintenance washing, either chlorine or acid can be injected into the backwash tank when a CEB (chemically enhanced backwash) is required. A CEB is conducted when required depending on the membranes' fouling condition based on prevailing water quality characteristics. The typical frequency for this operation is once a day to once a week. CEB involves soaking the membranes in solutions of sodium hypochlorite or acid for a desired period of time. The soaking times are typically 15 to 45 minutes per chemical and are operator adjustable. CEB can enhance the permeability recovery and extend the time interval between full chemical cleanings.

If foulant has accumulated and TMP baseline reaches approximately 15 psi, a CIP process (full chemical cleaning) is recommended in order to restore the membrane permeability. The CIP process consists of recirculation and soaking by two different chemicals: citric acid followed by sodium hypochlorite. A typical procedure is one hour circulation followed by 4 to 8 hours soaking. Those times vary depending on the membranes' fouling condition.

Typical parameters for the CM-3 ceramic membrane pilot are given in Table 4-2.

Table 4-2: Typical CM-3 Membrane System Pilot Test Parameters

Parameter	Value
Number of Modules	1
Filtration Period, minutes	30-90
Filtrate Flux, gfd	50-200
Maximum Trans-Membrane Pressure (psi)	35
Backwash Process	Reverse flow/ flushing
Backwash Cycle Time, minutes	1
Backwash Flow, gpm/module	50-250 gpm

Table 4-2: Typical CM-3 Membrane System Pilot Test Parameters (cont'd)

Parameter	Value
Filtrate Waste due to Backwash, gal/module	13
Backwash Frequency (# per day)	16-48
Total Waste Volume, gpd	200-625
CIP Procedure (Sodium Hypochlorite)	
Recirculation Duration, minutes	120-240
Recirculation Flow, gallons/minute per module	3-5 gpm
No. of Backwashes Before Filtration Resumes	1-2
Total Duration (with Backwashes), minutes	300
Total (CIP) Waste Volume, gallons per module	25
Concentration, ppm	1000-3000
Temperature, °F (°C)	80-100 (27-38)
CIP Procedure (Citric Acid)	
Recirculation Duration, minutes	120
Recirculation Flow, gallons/minute per module	3-5 gpm
No. of Backwashes Before Filtration Resumes	1-2
Total Duration(Including Backwashes), minutes	300
Total Waste Volume, gallons per module	25
Concentration, % on weight-to-weight basis	1
Temperature, °F (°C)	80-100 (27-38)
Integrity Test	
Frequency	Subject to local requirements
Purge Duration, minutes	2
Pressure Hold Time, minutes	10
Decay Duration, minutes	10

4.3 System requirements

The CM-3 pilot system requires 230v, 3ph power @ 50A and comes with a 460v – 230v step down transformer.

Membrane surface area: 269 ft^2; flux at 100 gfd = 18.7 gpm flow rate

The system is designed for indoor use only.

Oil free compressed air or use supplied air compressor unit, 5.6 scfm @ 120 psi

5 Operation and Results

The complete pilot test (i.e., includes design runs 1, 2 and 3) was conducted from August 14 to December 14, 2017. Each design run was conducted for 28 days and was preceded by an optimization period with a CIP performed before and after each run. Prior to design run 1, a test was performed on clean water, and the TMP was recorded at various fluxes, as shown in Table 5-1.

Table 5-1: Clean Water Flux Test Results

Flux (gfd)	Flow rate (gpm)	TMP (psi)
50	9.3	1.2
75	14	1.8
100	18.7	2.2

The three design runs were operated at the following fluxes: a) design run 1 and +2 at 50 gfd, and design run 3 at 150 gfd. Chart 5.1 below shows the flux comparison between runs, and Table 5-1a gives a comparison between the TMPs at the start and end of each CEB for all three runs.

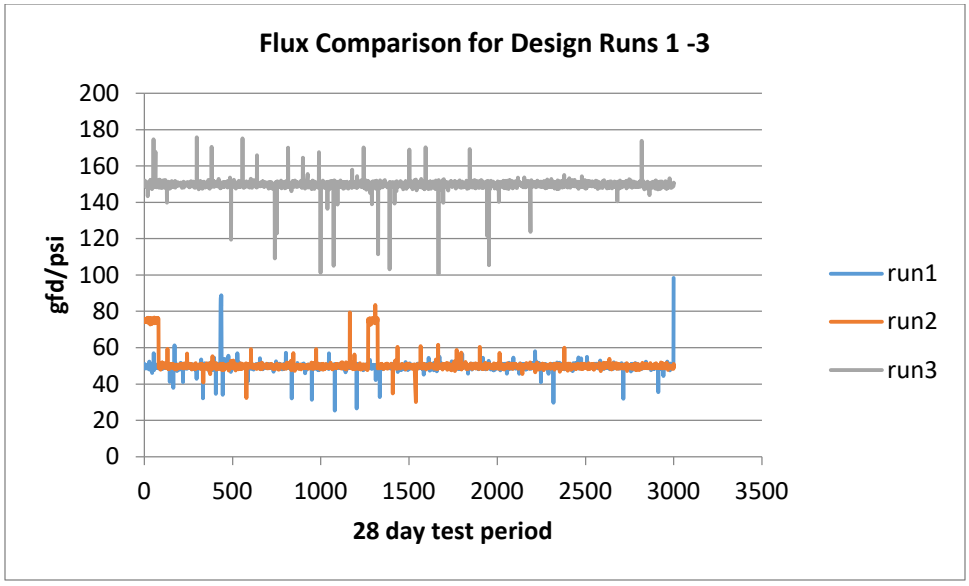


Chart 5.1: Flux Comparison Between Design Runs 1-3

Table 5-1a: TMP Comparison Between CEBs for Design Runs 1-3

	Design Run 1	Design Run 2	Design Run 3
Average TMP After CEB	3.1 psi	2.3 psi	6.9 psi
Average TMP Prior to Next CEB	11.7 psi	4.4 psi	12.0 psi
Average TMP Rise Between CEBs	8.6 psi	2.1 psi	5.1 psi

5.1 CIP Results

A CIP was conducted at the end of each design run. The CIP recipe used for all three runs is as follows (more details in Appendix A):

- solutions of 1% citric acid and 3000 ppm of sodium hypochlorite were used independently for cleaning;
- The chemical solution was prepared at ambient temperature before recirculation.
- Each chemical was recirculated for 2 hours followed by 2 hours of soaking.
- The Acid CIP solution was heated to approximately 80 deg. F for design runs 1 and 2

- The Hypochlorite CIP solution was heated to approximately 80 degrees Farenheit for design runs 1 and 2.

Following each chemical, the unit was placed back into service at various fluxes and the TMPs were recorded and compared to verify that this cleaning regime was indeed removing all of the foulant material from the membrane. The following charts and tables show the results of these tests; note that source water – not clean water – was used for each of these tests, so the resultant TMPs are 33-36% higher than the Clean Water Flux Test Results recorded in Table 5-1.

5.1.1 Run 1 CIP results

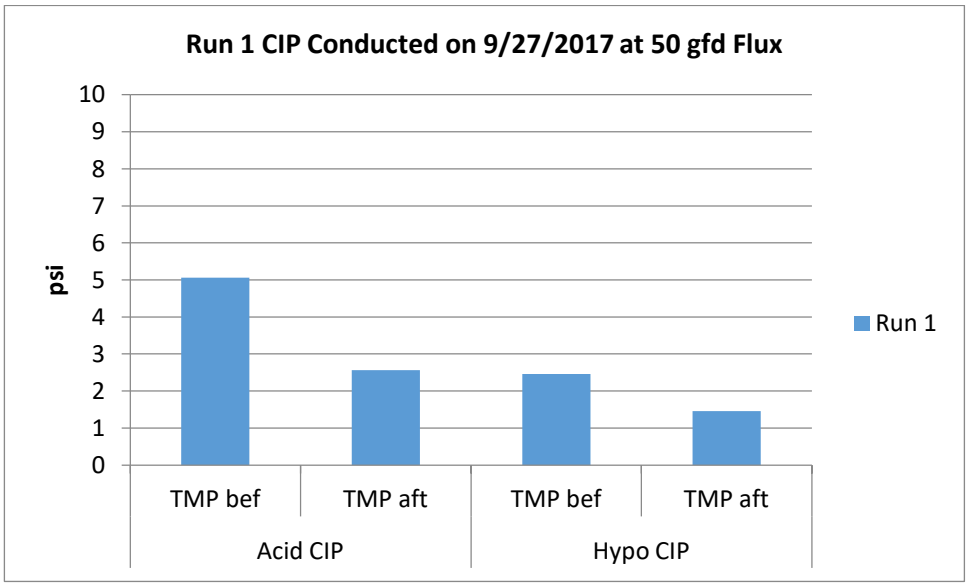


Chart 5.1a: Run1 CIP Conducted Using TID Water

5.1.2 Run 2 CIP results

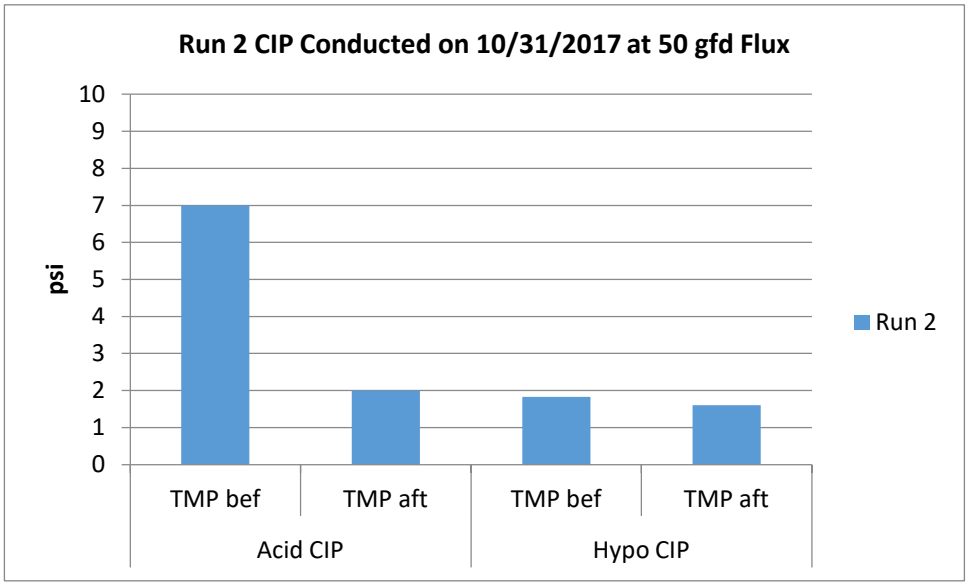


Chart 5.1b: Run 2 CIP Conducted Using Reeder Reservoir Water

Table 5-2: Run 2 Post Acid CIP Results at Various Fluxes Using Reeder Reservoir Water

Flux (gfd)	Flow rate (gpm)	TMP (psi)
50	9.3	2.0
75	14	2.7
100	18.7	3.6
125	23.3	4.3

Table 5-3: Run 2 Post Hypo CIP Results at Various Fluxes Using Reeder Reservoir Water

Flux (gfd)	Flow rate (gpm)	TMP (psi)
50	9.3	1.6
75	14	2.4
100	18.7	3
125	23.3	3.9

5.1.3 Run 3 CIP results

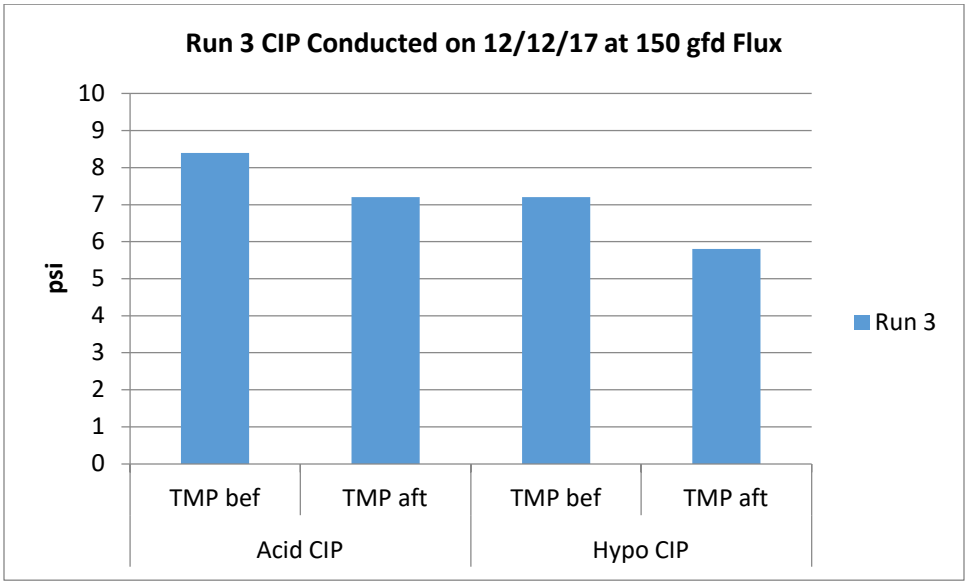


Chart 5.1c: Run 3 CIP Conducted Using Reeder Reservoir Water

Table 5-4: Run 3 Post Acid CIP Results at Various Fluxes on Reeder Reservoir Water

Flux (gfd)	Flow rate (gpm)	TMP (psi)
50	9.3	2
75	14	2.7
100	18.7	3.6
125	23.3	4.3
150	28	7.2

Table 5-5: Run 3 Post Hypo CIP Results at Various Fluxes on Reeder Reservoir Water

Flux (gfd)	Flow rate (gpm)	TMP (psi)
50	9.3	1.6
75	14	2.4
100	18.7	3
125	23.3	3.9
150	28	5.8

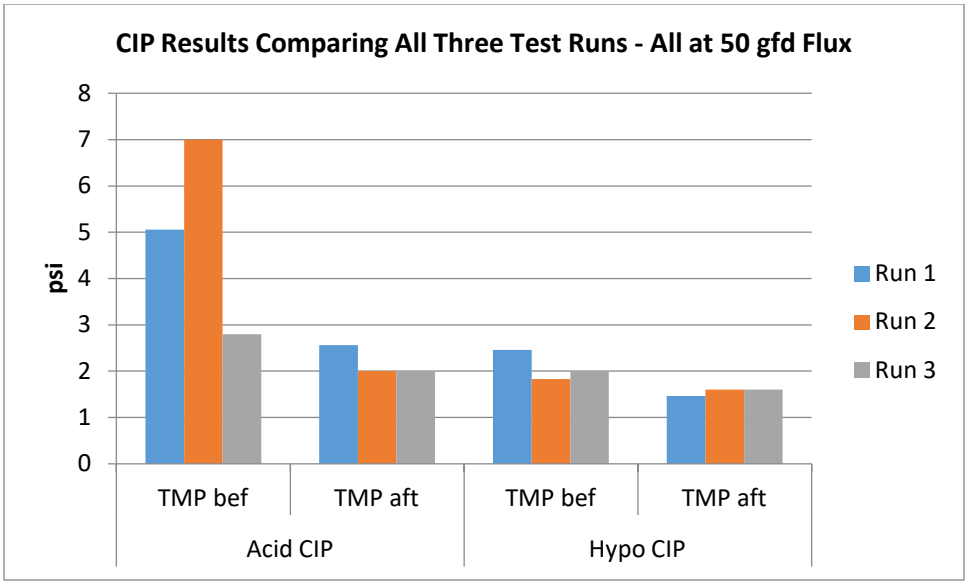


Chart 5.1d: Comparison of CIP Results for All Three Runs

Since there are no test results following the initial CIP (prior to run 1) – only a clean water test – the 1.4 psi TMP after the hypochlorite CIP following run 1 (shown as the blue bar on the right in Chart 5.1d above) must be compared to the 2.0 psi TMP baseline recorded during run 1. The CIP after run 1 actually lowered the TMP. Chart 5.1d also shows that the TMPs following runs 2 and 3 – both on Reeder Reservoir water – were exactly the same (1.6 psi), indicating that the CIPs were able to remove all of the foulant on the membrane.

5.2 Design run 1

Design run 1 was conducted from 08/30/17 to 9/27/17. This run was conducted to evaluate the ceramic membrane performance using water from the Talent Irrigation District (TID). The operational settings established during the optimization period prior to starting design run 1 are shown in Table 5-6 below.

Table 5-6 Operating Conditions Established During Design Run 1

Parameter	Value
Flux	50 gfd
Flow rate	9.3 gpm
Filtration / Backwash (BW) cycle	45 minutes
CEB cycle	1x /day
CEB soak time	60 minutes
Coagulant dose (ppm 50% ACH)	2 ppm to CM3 unit, starting 9/17/17

Feed water was fed into the CM3 unit's mix tank and kept at a low tank volume (i.e., approximately 50 gallons). An overflow was installed on the tank to ensure this level was maintained. Coagulant at 8 ppm (as 50% ACH) was intended to be injected by the Owner into the raw feed supply at the start of the test run (common for all MFEMS), but it was later discovered during design run 2 that no ACH had been injected during run 1 or during run 2 prior to 10/11/2017. However, due to elevated TMP levels observed during run 1, , AASI began the injection of 2 ppm coagulant (as 50% ACH) into the CM-3 pilot mix tank to help control TMP levels; this can be observed in the data on Chart 5.2.1a.

Starting TMP value was 3.5 psi (at 50 gfd flux) with peaks (i.e., highest TMP value between backwash or CEB events) ranging from 10 psi to 20 psi (avg) at the end of the 45-minute BW cycle and baseline TMP 2-4 psi. After adding the 2 ppm ACH dose, TMP peaks were reduced to 10–15 psi with baseline (i.e., post-backwash) TMP between 2-3 psi, recovered after CEB events. The following charts, 5.2a and 5.2b, show the impact of adding the 2 ppm ACH dose. The general trend shown in chart 5.2a for TMP peaks is sharply upward with slope of 1.1 (roughly) whereas, with the 2 ppm ACH dose (refer to Chart 5.2b), the general trend for TMP peaks is moderately upward with slope of 0.13 (roughly).

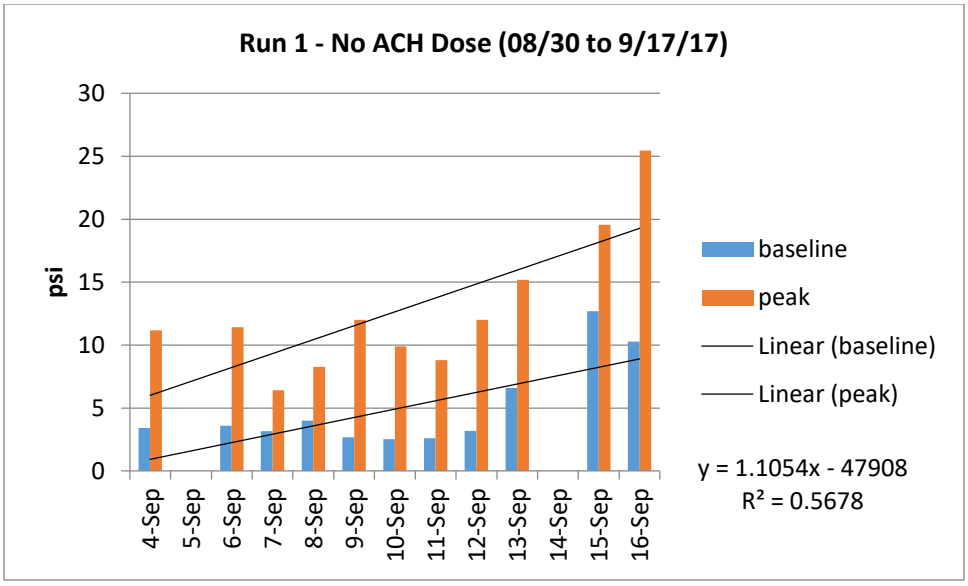


Chart 5.2a: TMP Baseline - Peak Analysis with No ACH Dose

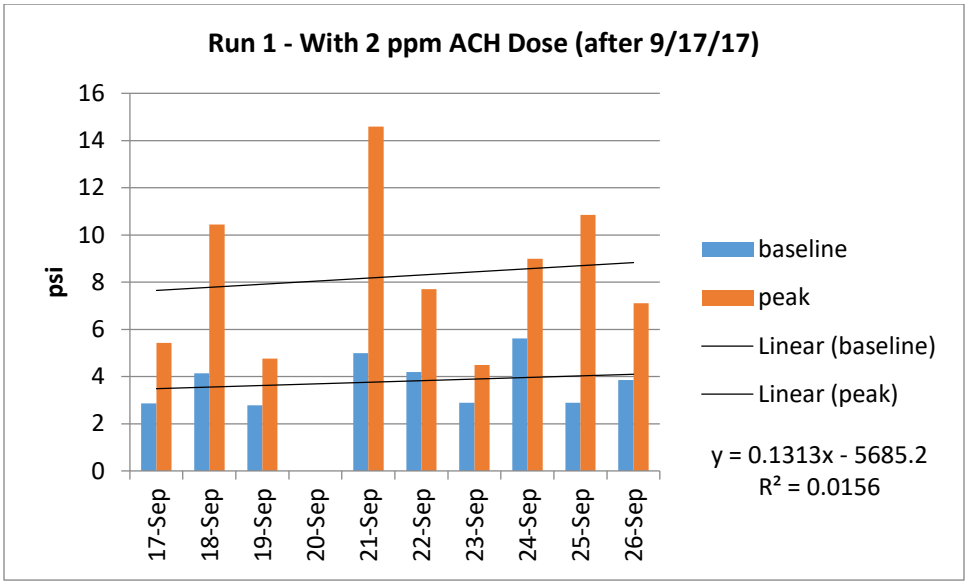
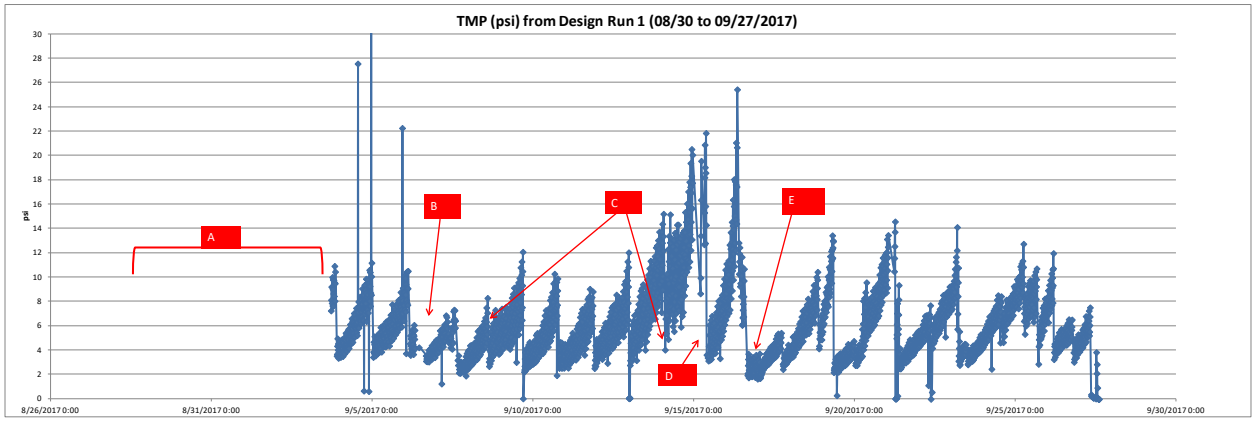


Chart 5.2b Baseline and Peak TMP Analysis with 2 ppm ACH Dose After 9/17/2017

Backwash events numbered 896 throughout the 28 day test which amounted to 11,827 gallons of backwash waste (this accounts for CEB waste volume as well). CEB events occurred 28 times (60 min total soak time / event) in total. Total filtrate volume (i.e., minus BW and CEB waste volumes) was 346,744 gallons. Total feed to system volume was 358,571 gallons. Total system recovery was 96.7 %.

Total reject rate (i.e., as a percentage of feed, BW+ CEB waste volumes) was 3.3 %. Lastly, a CIP was performed prior to starting and directly after the 30 day test. The results from the CIP following this test run are shown in section 5.1 on chart 5.1a.

5.2.1 Data Charts for Design Run 1



A Missing data due to data storage rollover (from 08/30 to 09/03/2017)
B Downtime due to troubleshooting for (PDT) air leak ~10hours
C CEB was ineffective here. This is indicative of a dosing problem (i.e., airlock or prime loss)
D Corrected CEB (hypo) dosing pump problem. Seems to have been air locked
E Started 2 ppm ACH supplement to feed at this point forward

Chart 5.2.1a: TMP Data for Design Run 1 on TID Water (psi)

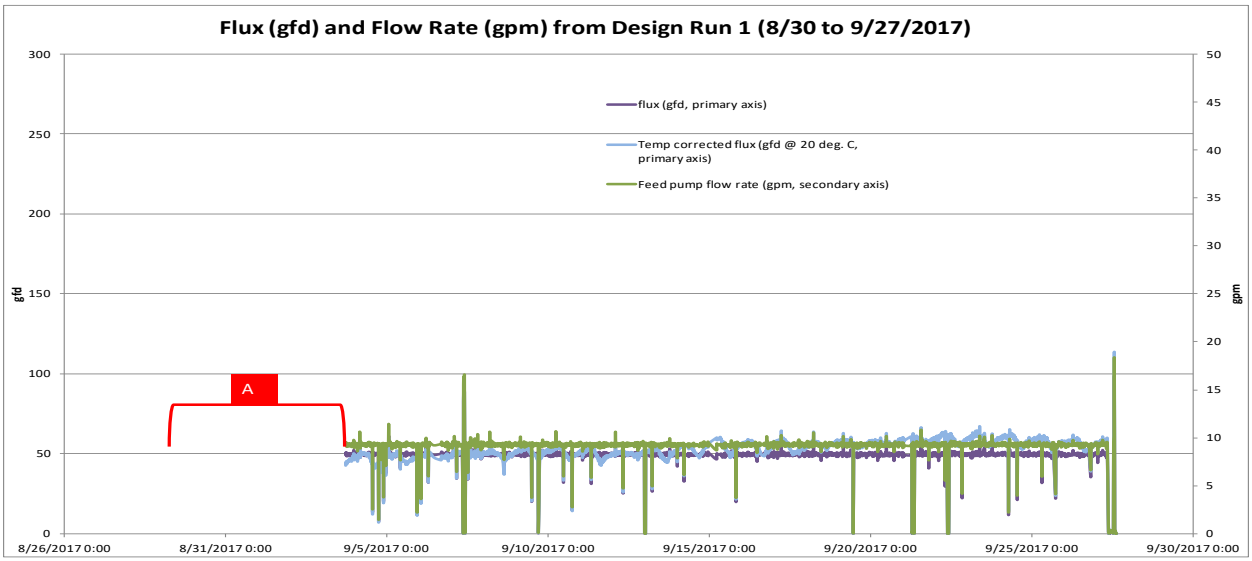
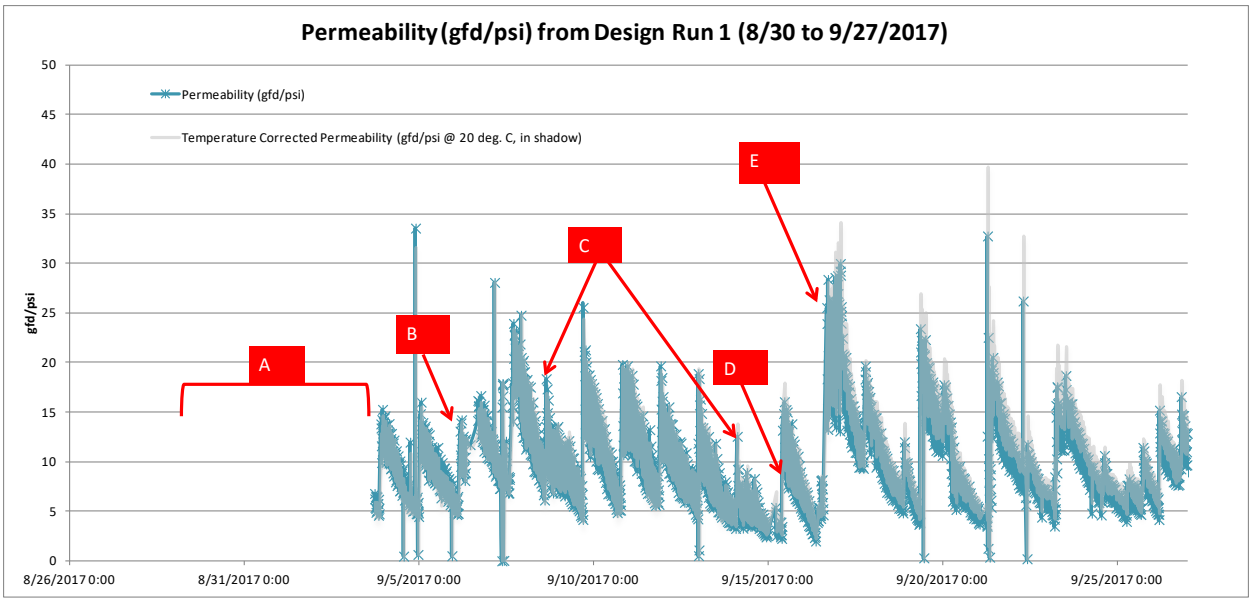


Chart 5.2.1b: Flux and Flow Rate for Design Run 1



A Missing data due to data storage rollover (from 08/30 to 09/03/2017)
B Downtime due to troubleshooting for (PDT) air leak ~10hours
C CEB was ineffective here. This is indicative of a dosing problem (i.e., airlock or prime loss)
D Corrected CEB (hypo) dosing pump problem. Seems to have been air locked
E Started 2 ppm ACH supplement to feed at this point forward

Chart 5.2.1c: Permeability for Design Run 1 (gfd/psi)

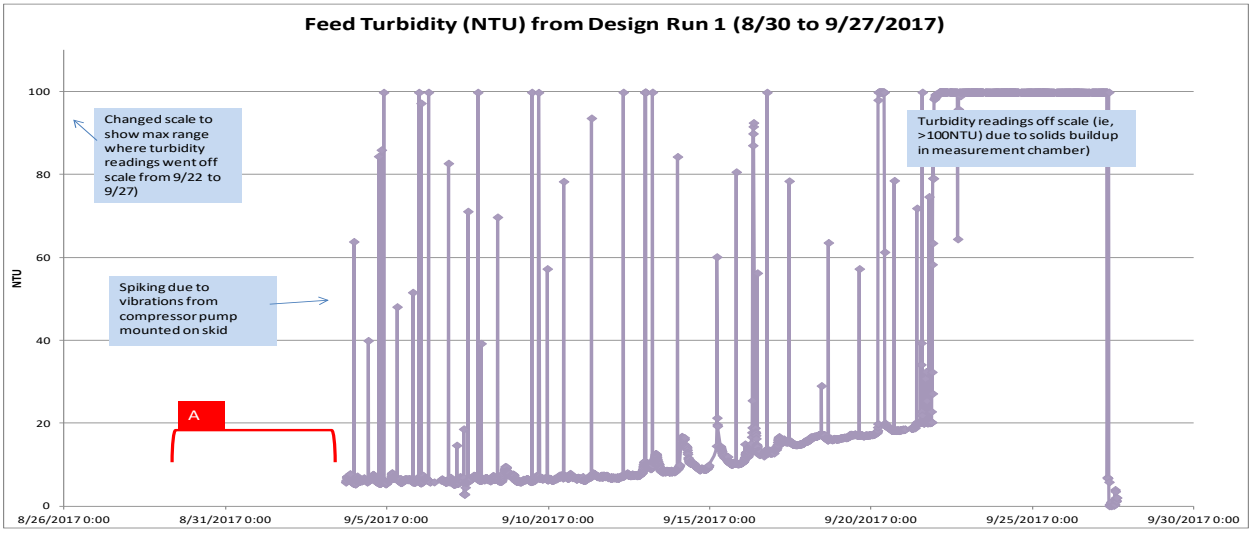


Chart 5.2.1d: Feed Turbidity for Design Run 1 (NTU)*

*Note: the drop off from 9/22 to 9/27 is a period where feed turbidity readings reached maximum range (i.e., 100 NTU) on the turbidimeter – due to solids buildup in measurement chamber. This was cleaned at end of run (9/27/2017).

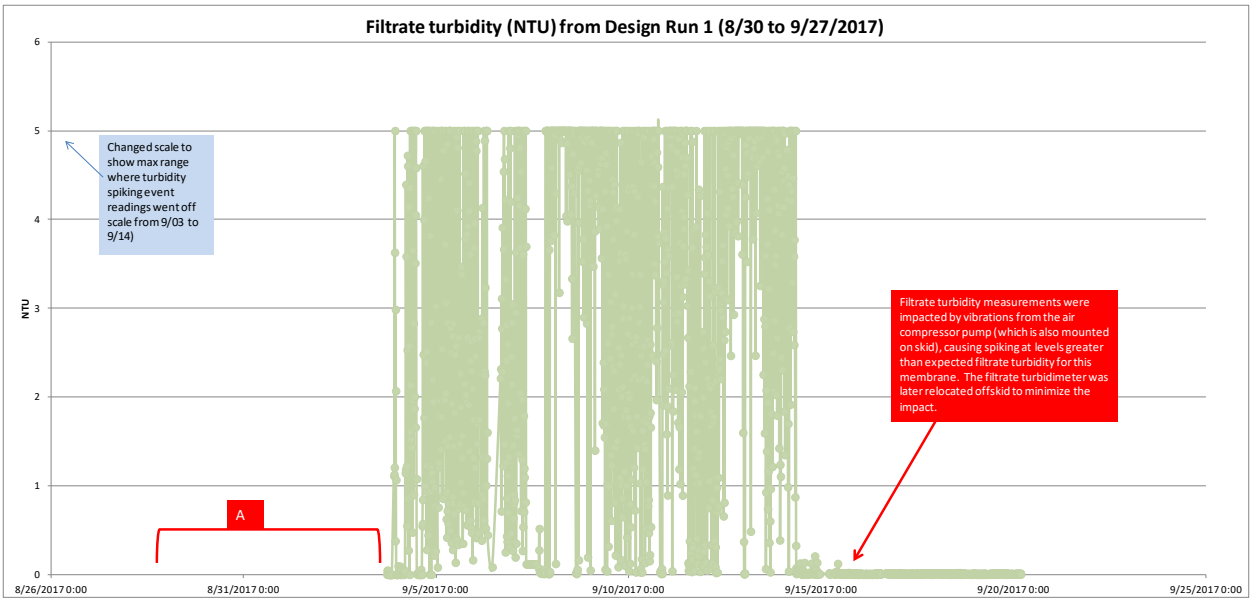


Chart 5.2.1e: Filtrate Turbidity for Design Run 1 (NTU)

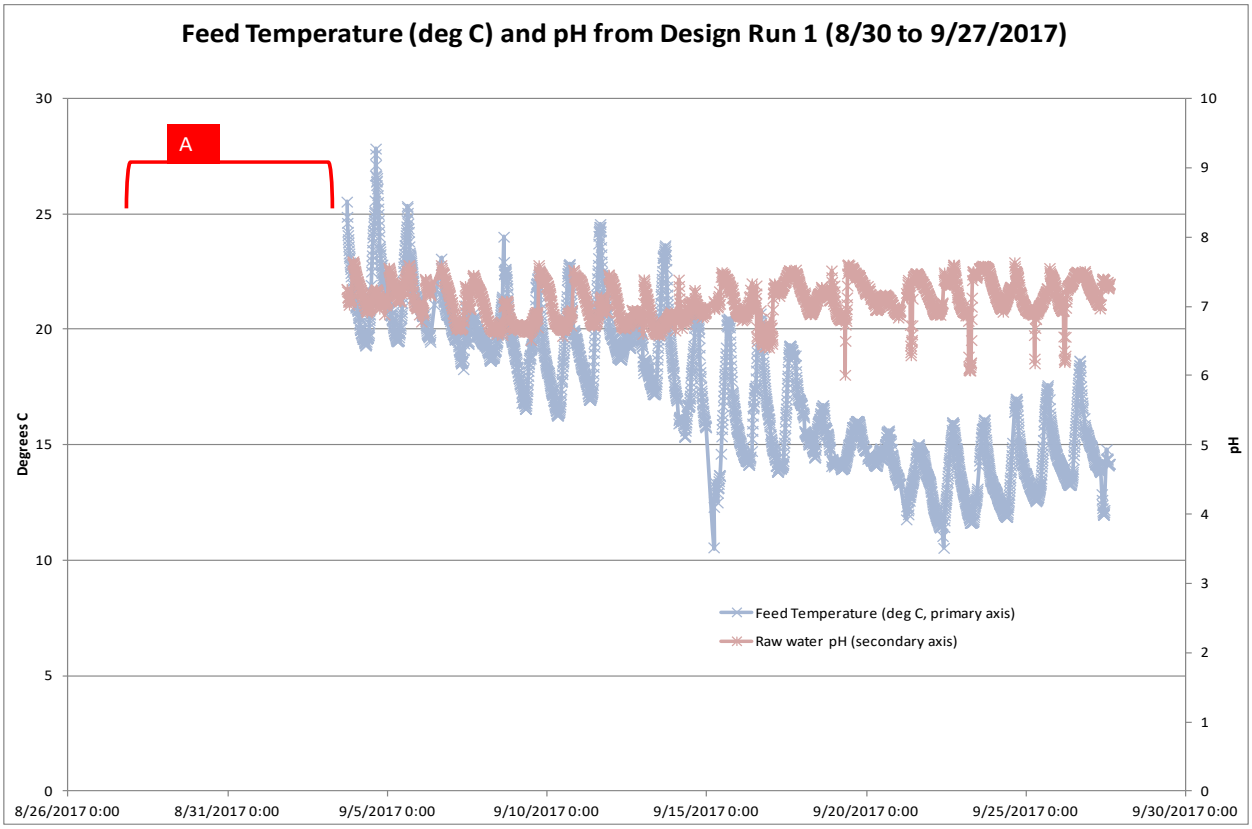


Chart 5.2.1f: Feed pH and Temperature for Design Run 1

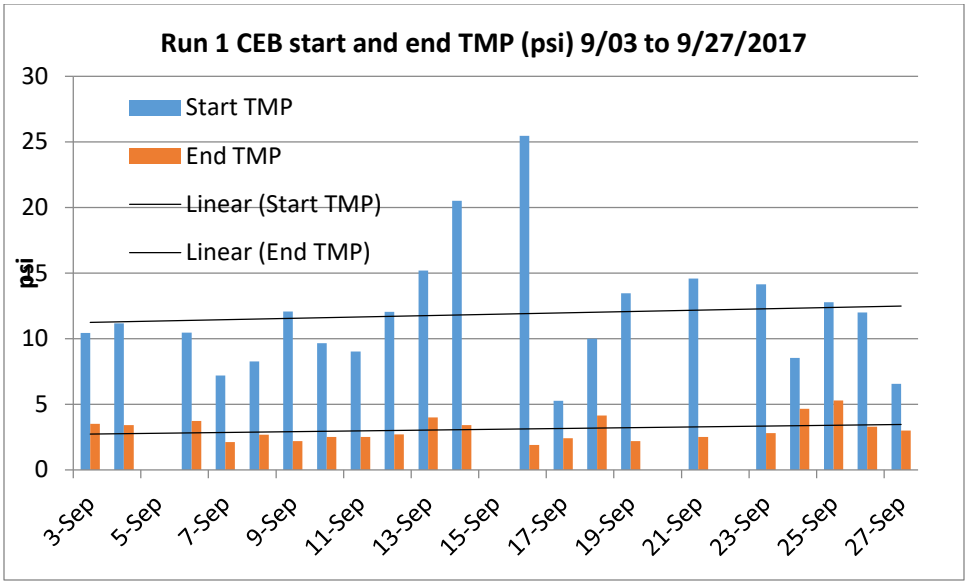


Chart 5.2.1g: CEB Results for Design Run 1

5.3 Design Run 2

Design run 2 was conducted from 10/03/17 to 10/31/17. This run was conducted to evaluate the ceramic membrane performance with a conservative flux using water from the Reeder Reservoir. The operational settings established during the optimization period prior to starting design run 2 are shown in Table 5-7.

Table 5-7 Operating Conditions Established During Design Run 2

Parameter	Value
Flux	50 gfd
Flow rate	9.3 gpm
Filtration / Backwash (BW) cycle	60 minutes*
CEB cycle	1x /day**
CEB soak time	60 minutes
Coagulant dose (as 50% ACH)	8 ppm in common feed, started on 10/11

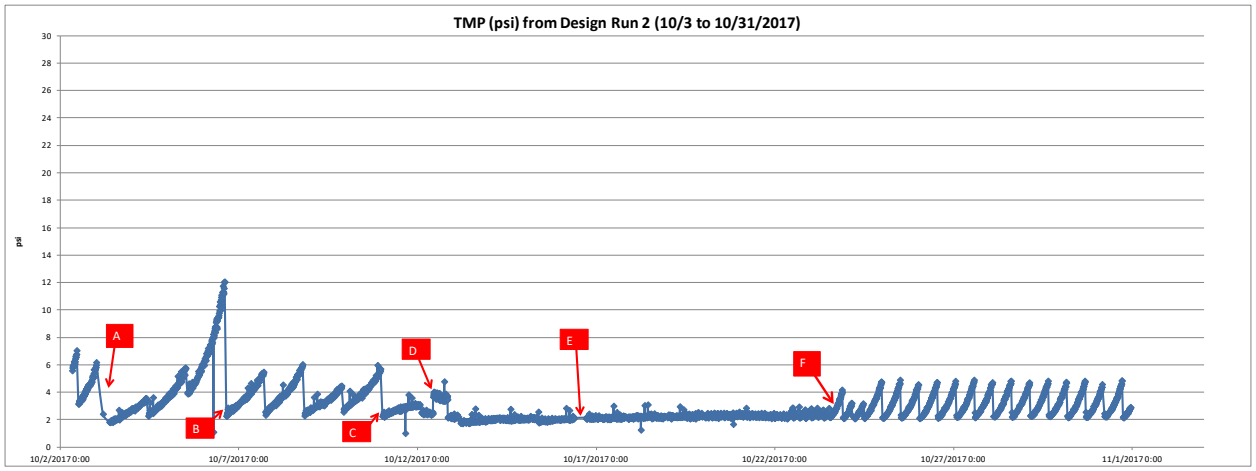
**Increased filtration /BW cycle time from 60 to 720 (12 hr) minutes on 10/23/17 to increase recovery*
***Decreased CEB frequency from 1x/day to 1x/week on 10/23/17*

Pretreated feed water was fed into the CM3 unit's mix tank in the same way as done for design run 1. However, the feed water source for design run 2 was changed to the Reeder Reservoir. On 10/11/17, it was discovered that the

coagulant was not being injected into the raw feed supply; following this date, 8 ppm (as 50% ACH) was injected in the feed line common to all pilots.

Starting baseline TMP value was 1.8 psi (at 50 gfd flux). TMP peaks ranged from 2 – 12 psi. Backwash events numbered 496 throughout the 28 day test, which amounted to 6,547 gallons of backwash waste (this accounts for CEB waste volume as well). CEB events occurred 21 times (60 min total soak time / event) in total. Total filtrate volume (i.e., minus BW and CEB waste volumes) was 355,846 gallons. Total feed to system volume was 362,393 gallons. Total system recovery was 98.1%. Total reject rate was 1.8%. A CIP was conducted upon completion of this test run; the results are given in section 5.1.

5.2.1 Data Charts for Design Run 2



(A) Test phase 2 start at 50 gfd flux (9.3 gpm flow rate). 60 min BW cycle, 1x/day CEB
(B) Increased Hypochlorite CEB dose from 100 ppm to 300 ppm
(C) Starting point at which 8 ppm ACH was injected to raw feed water
(D) Increased flux to 75 gfd, directly after phone meeting with Keller. This was started at 10:30 am and ended later in the same day at 6:30pm, where flux was returned to 50 gfd per Keller's request.
(E) System shutdown due to apparent program hang-up at time of a CEB event. Start at 9:20 am and ended at 3:30pm
(F) Increased filtration time (BW cycle) from 60 to 720 minutes and decreased CEB from 1x/day to 1x/week

Chart 5.3.1a: TMP Data for Design Run 2 (psi)

Notes to Key Box:

(B) - **We noticed that Hypo CEB was not cleaning effectively as shown by baseline TMP (i.e., post backwash TMP) between (A) and (B) points on chart 5.3.1a. TMP was climbing and not returning close to baseline as expected. After increasing to 300 ppm (from 100), we found the post BW TMP returning close to baseline as shown directly after point (B) on chart 5.3.1a. We believe the dose can be further optimized.**

(C) The 8 ppm dose was Owner fed/supplied as common to all MFEM systems. We did not add any supplemental coagulant during Run 2.

(D) During regular phone call with Bryan Phinney (Keller and Associates), it was discussed that since we started Run 2 expecting the Owner to inject the common 8 ppm ACH dose , and later discovered it hadn't been dosed at all until 10/11/17, we would be permitted to increase flux to 75 gfd (i.e., from 50 gfd). It was increased on 10/12 at 10:30am. The driver for this request was the performance increase as shown on chart 5.2.1a from 10/11 – 10/31/17. Later the same day, I received a call from Bryan notifying that the flux change cannot occur and to revert back to the original starting flux of 50 gfd, which was promptly done on 10/12 at 8:20pm

(F) The changes made during this time period were done to decrease operating costs, in lieu of increasing flux, which was not permitted. Instead, we maximized operating conditions – i.e., increased filtration time / decreased BW frequency and decreased CEB frequency to cut waste and associated costs. The two changes (i.e., filtration time and CEB frequency) did not occur at the same time. A full day of data transpired before finding that decreasing CEB frequency from 1x/day to 1x/week was effective.

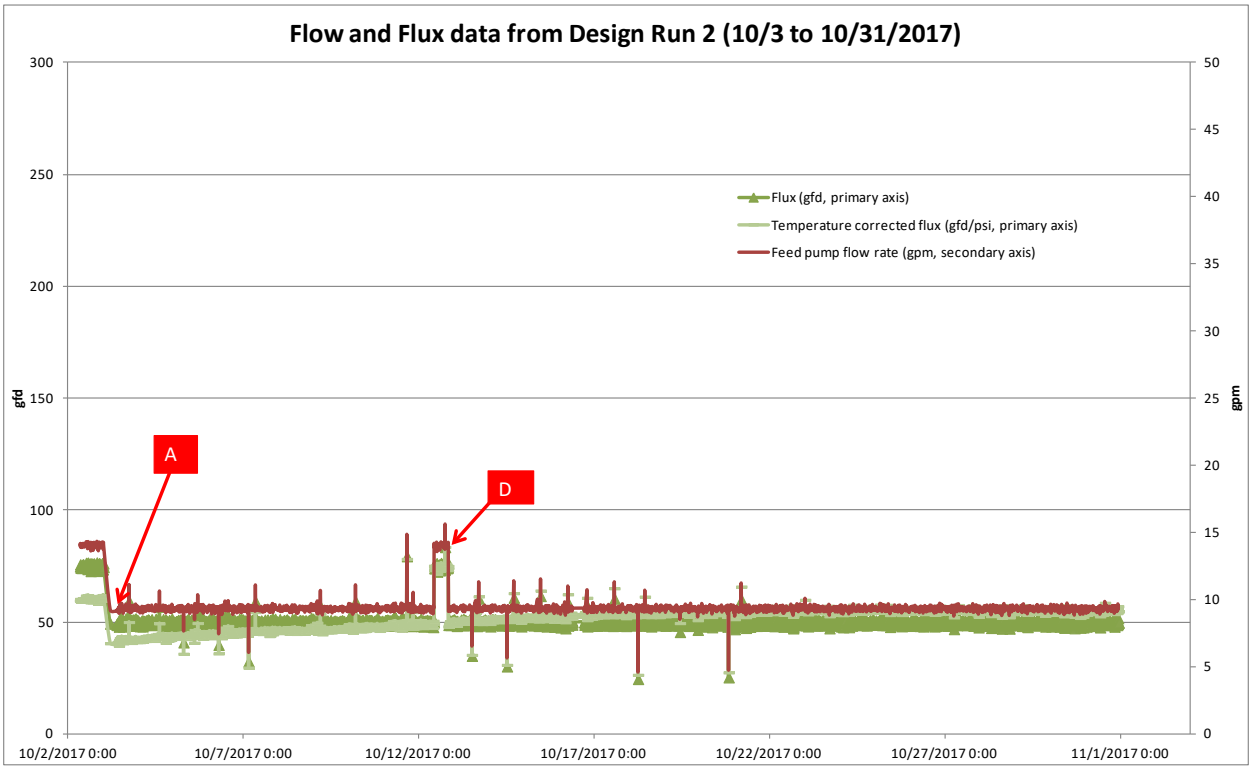


Chart 5.3.1b: Flow and Flux Data for Design Run 2*

*Note: refer to key box notes from chart 5.3.1a

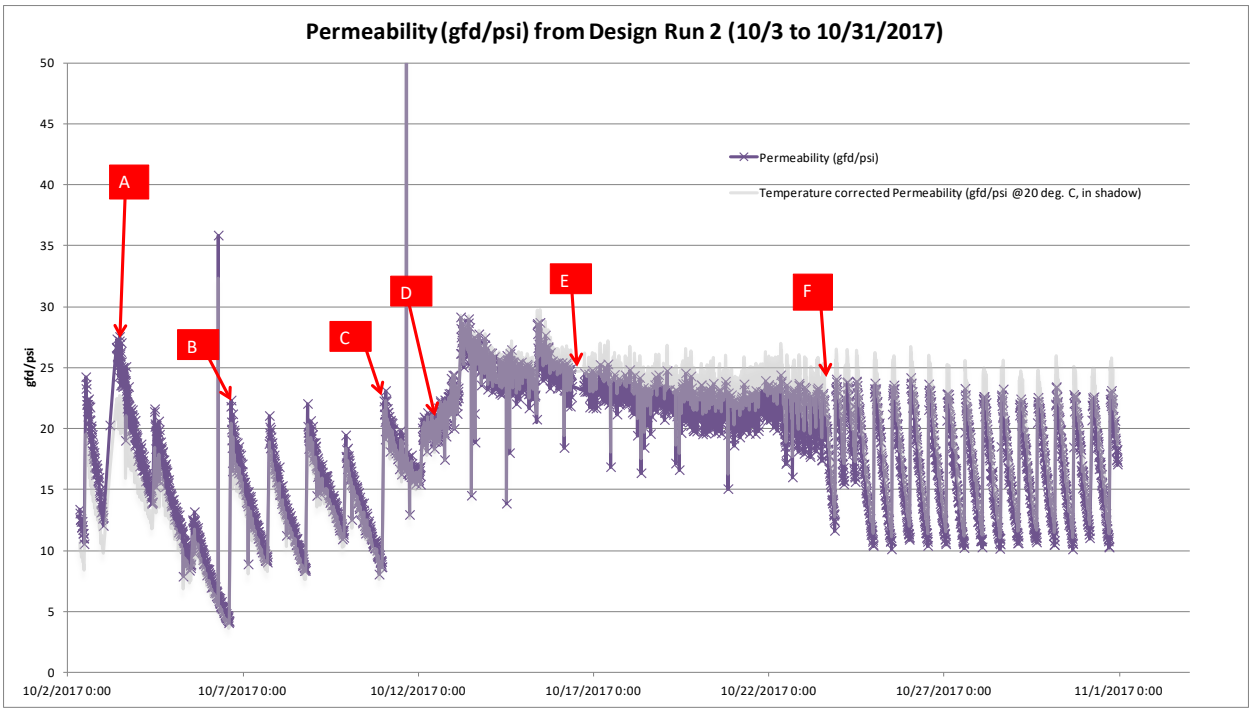


Chart 5.3.1c: Permeability Data for Design Run 2 (gfd/psi)*

*Note: refer to key box notes from chart 5.3.1a

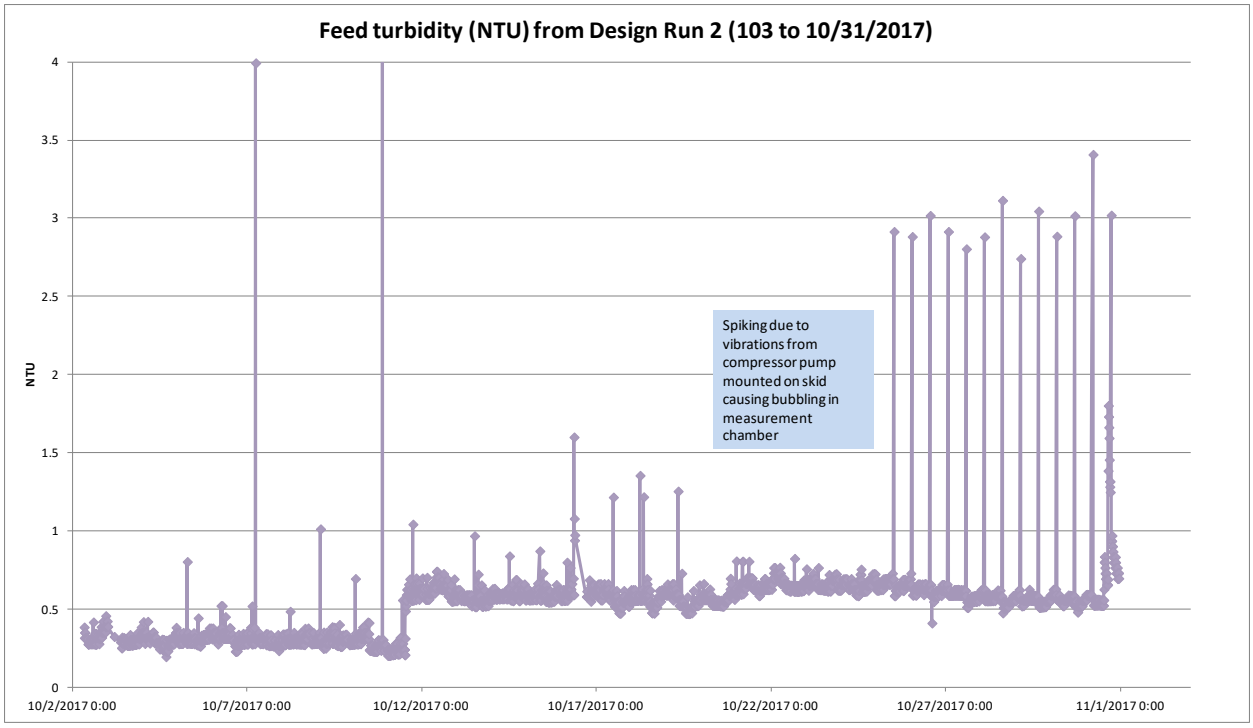


Chart 5.3.1d: Feed Turbidity Data for Design Run 2

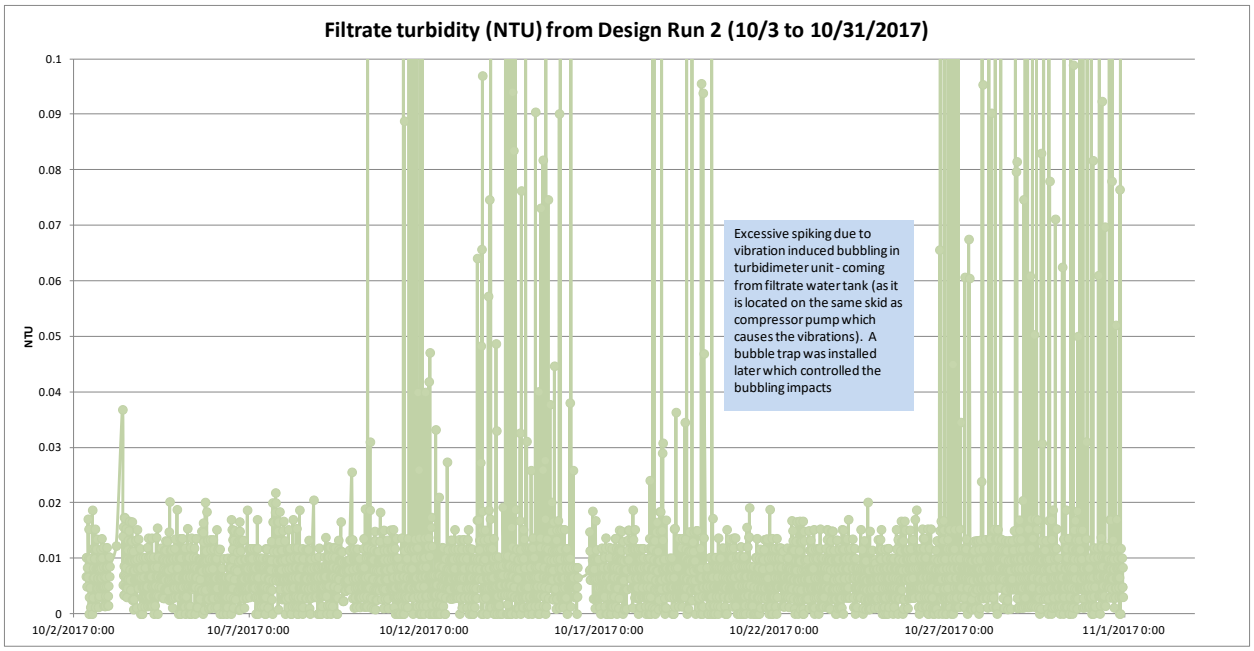


Chart 5.3.1e: Filtrate Turbidity Data for Design Run 2

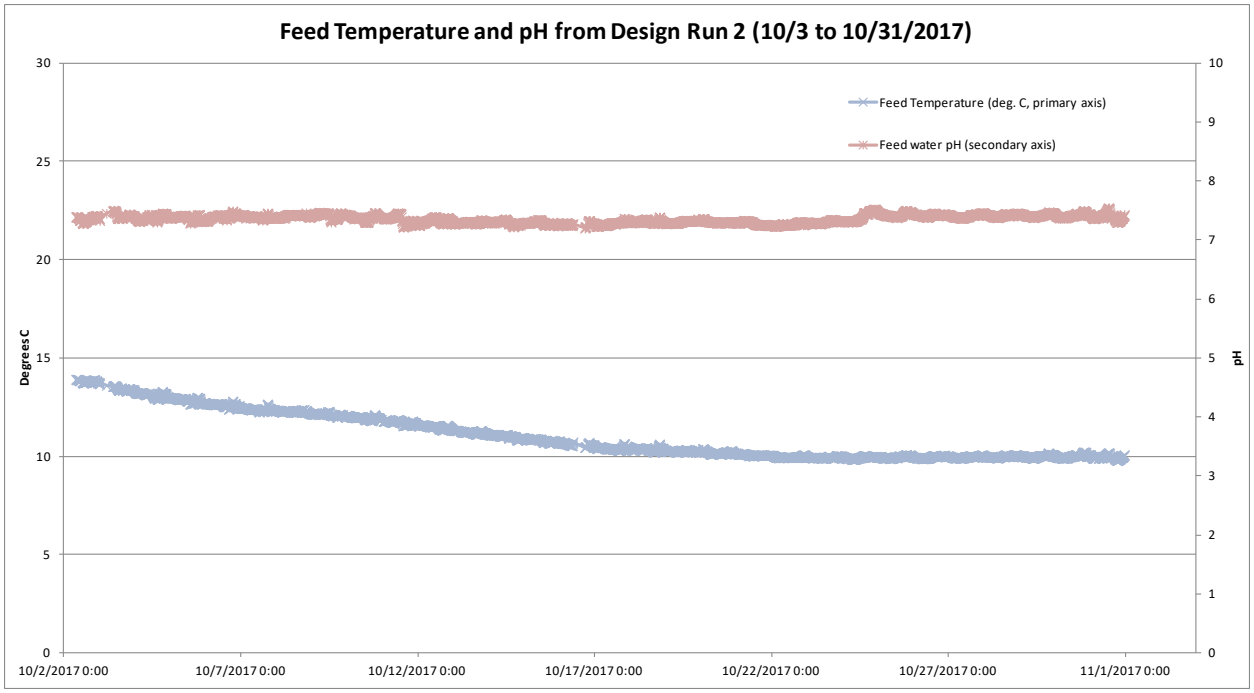


Chart 5.3.1f: Feed Temperature and pH Data for Design Run 2

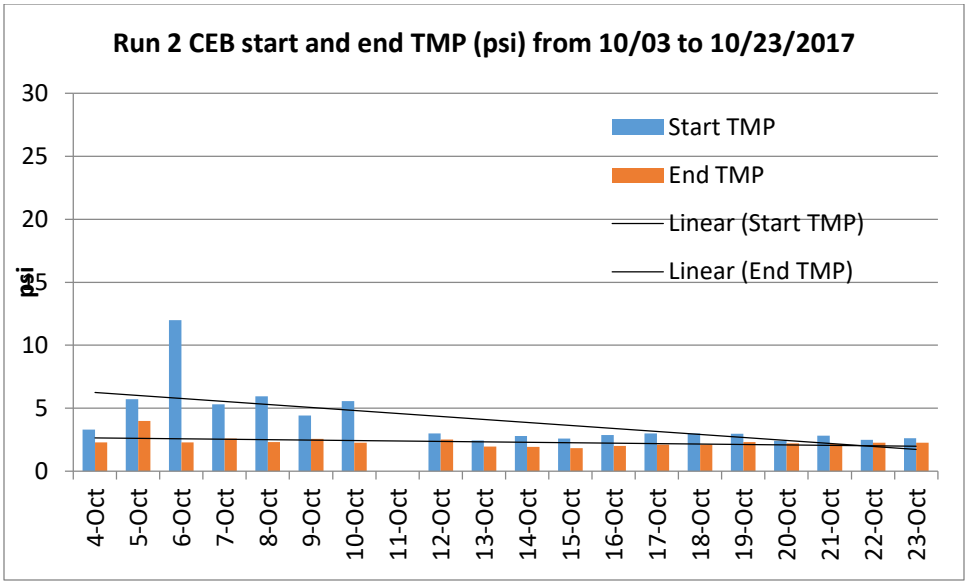


Chart 5.3.1g: CEB results for Design Run 2

5.4 Design run 3

Design run 3 was conducted from November 14 to December 12, 2017. Feed water from the Reeder Reservoir was fed into the CM3 unit mix tank and then pumped into the ceramic membrane system at a flow rate of 28 gpm (150 gfd). Operating conditions established during the optimization period for design run 3 are summarized below in Table 5-8.

Table 5-8 Operating Conditions Established During Design Run 3

Parameter	Value
Flux	150 gfd
Flow rate	28 gpm
Filtration / Backwash (BW) cycle	90 minutes
CEB cycle	1x /3days
CEB soak time	60 minutes
Coagulant dose (as 50% ACH)	8 ppm in common feed

Starting TMP baseline was 4.9 psi, and ending TMP baseline was 8.3 psi. TMP baseline (i.e., Post backwash TMP) average rate of increase / day was 0.15 psi. TMP peaks ranged from 9.4–17 psi (*peaks greater than 15 psi spiked briefly when coagulant feed was stopped from 11/25 to 11/27/2017*). Total system recovery

was 99.5%. Total Feed flow was 1,076,740 gallons. Total waste volume (i.e., from BW and CEB events) was 5,913.6 gallons. Total filtrate volume (i.e., minus waste volume) was 1,070,826 gallons. Total reject rate was 0.5%. A CIP was performed after this run; the results are given in Tables 5-4 and 5-5.

5.4.1 Data charts for Design run 3

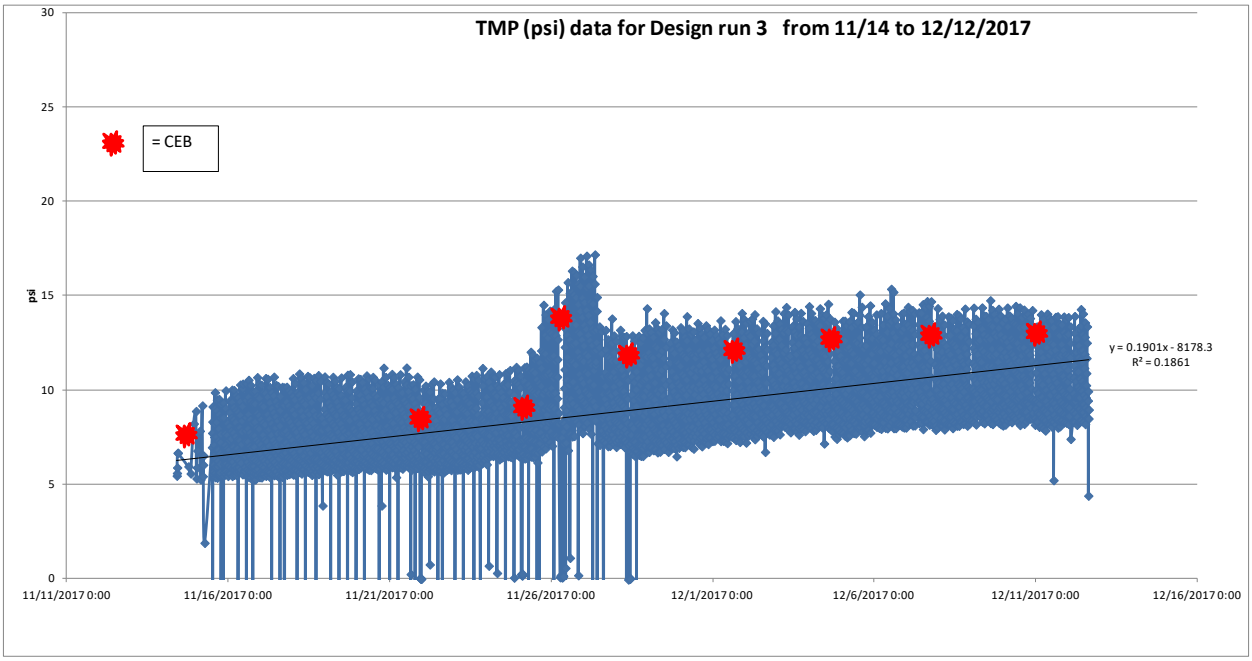


Chart 5.4.1a: TMP Data for Design Run 3 *

*Note: The TMP range between backwashing and CEB is higher on this run due to the higher flux /feed flow rate, and longer filtration time (i.e., 90 minutes) and as a result, higher solids loading.

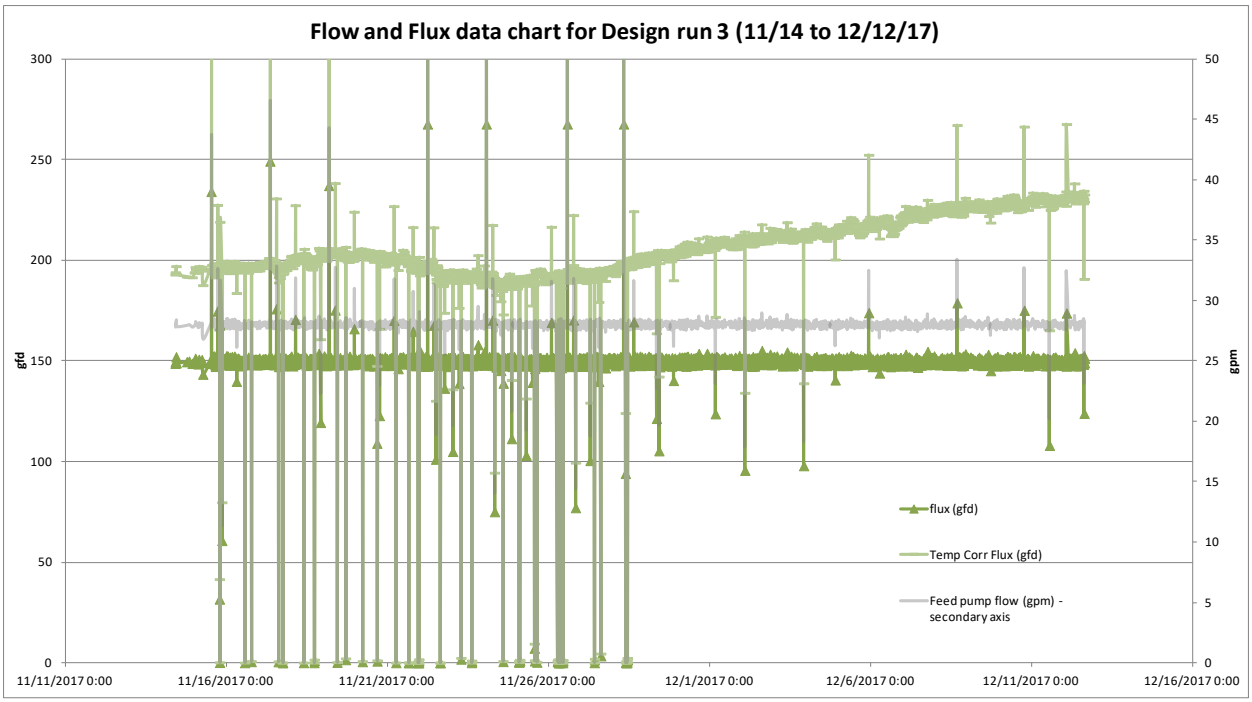


Chart 5.4.1b: Flow and Flux Data for Design Run 3

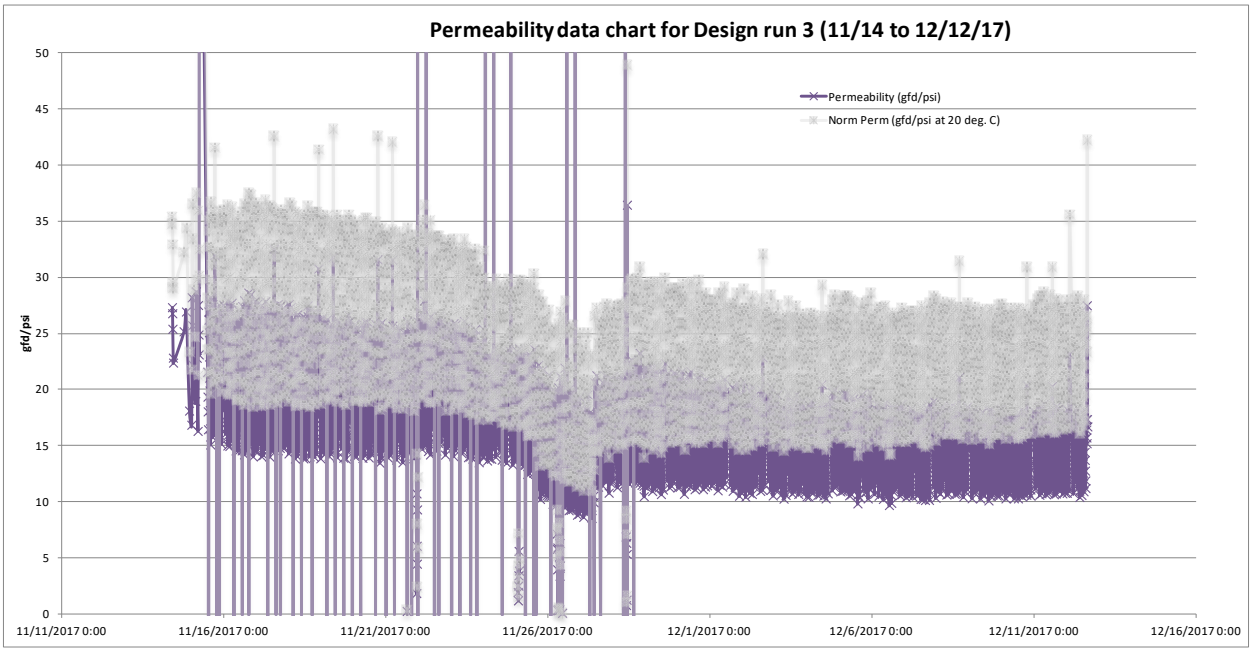


Chart 5.4.1c: Permeability Data for Design Run 3 (gfd/psi)

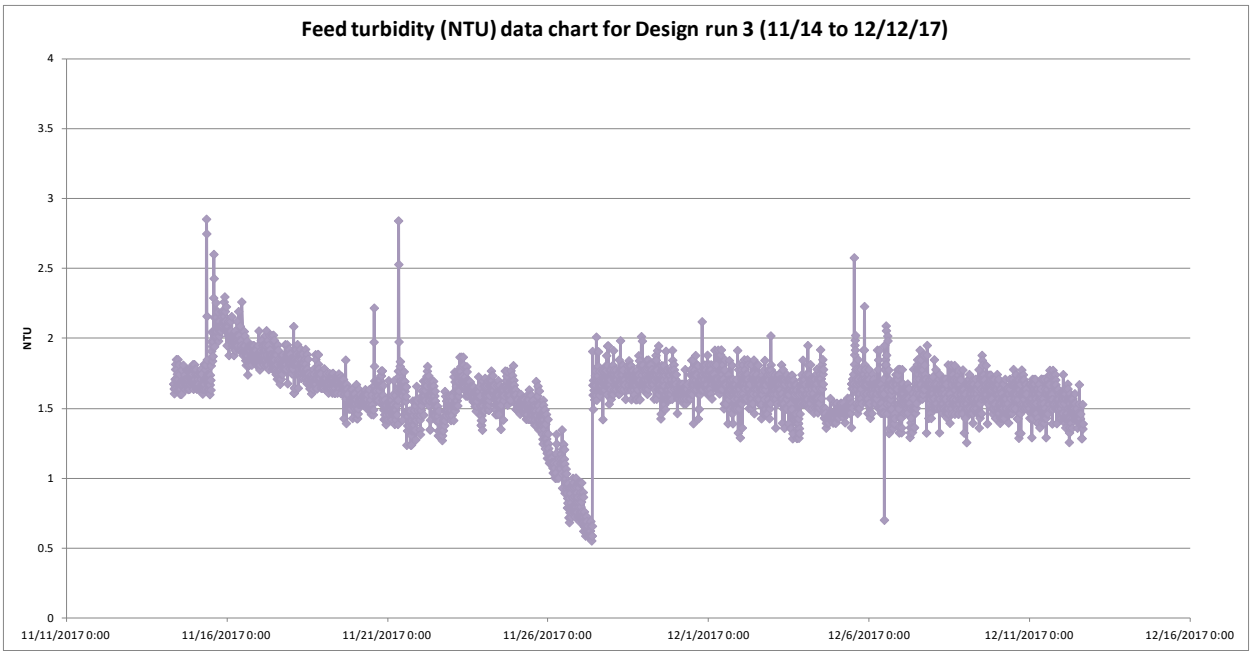


Chart 5.4.1d: Feed Turbidity Data for Design Run 3 (NTU)

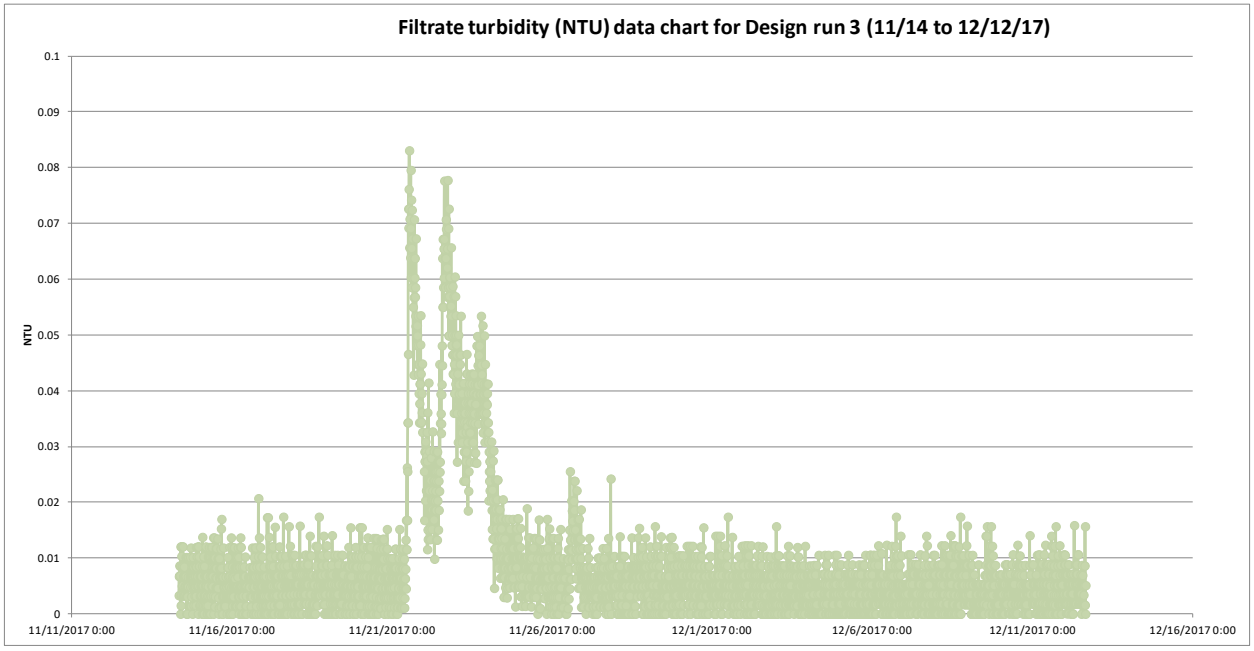


Chart 5.4.1e: Filtrate Turbidity Data for Design Run 3 (NTU)

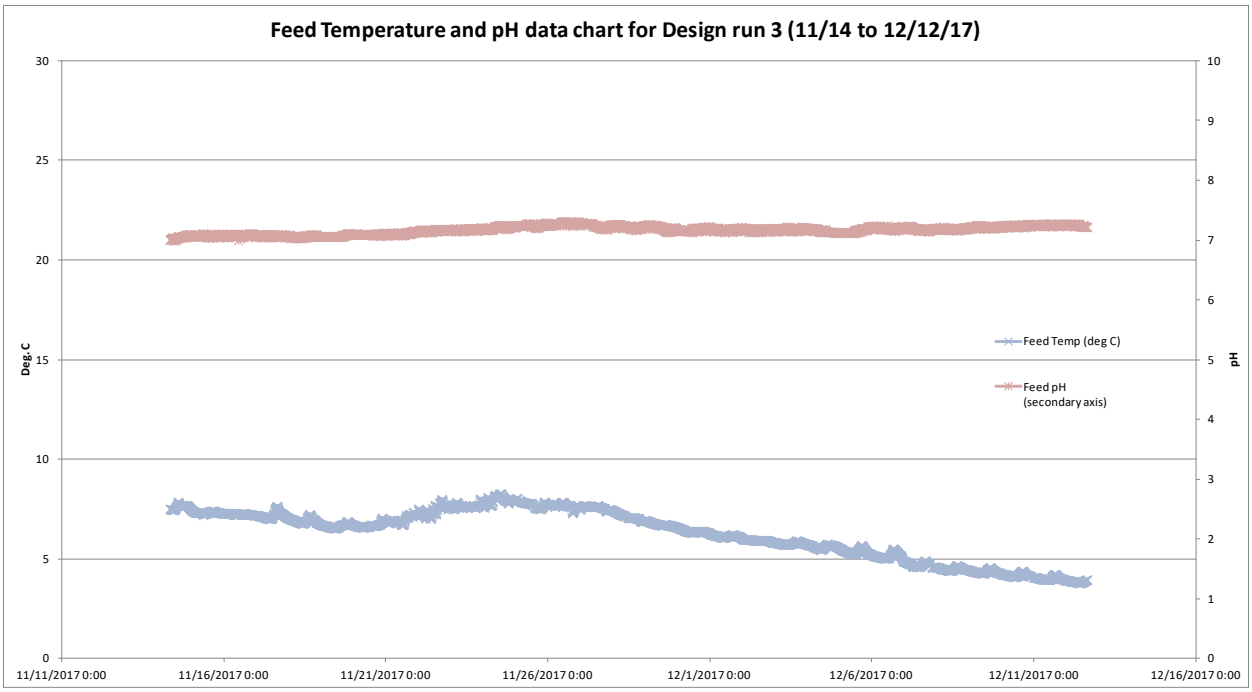
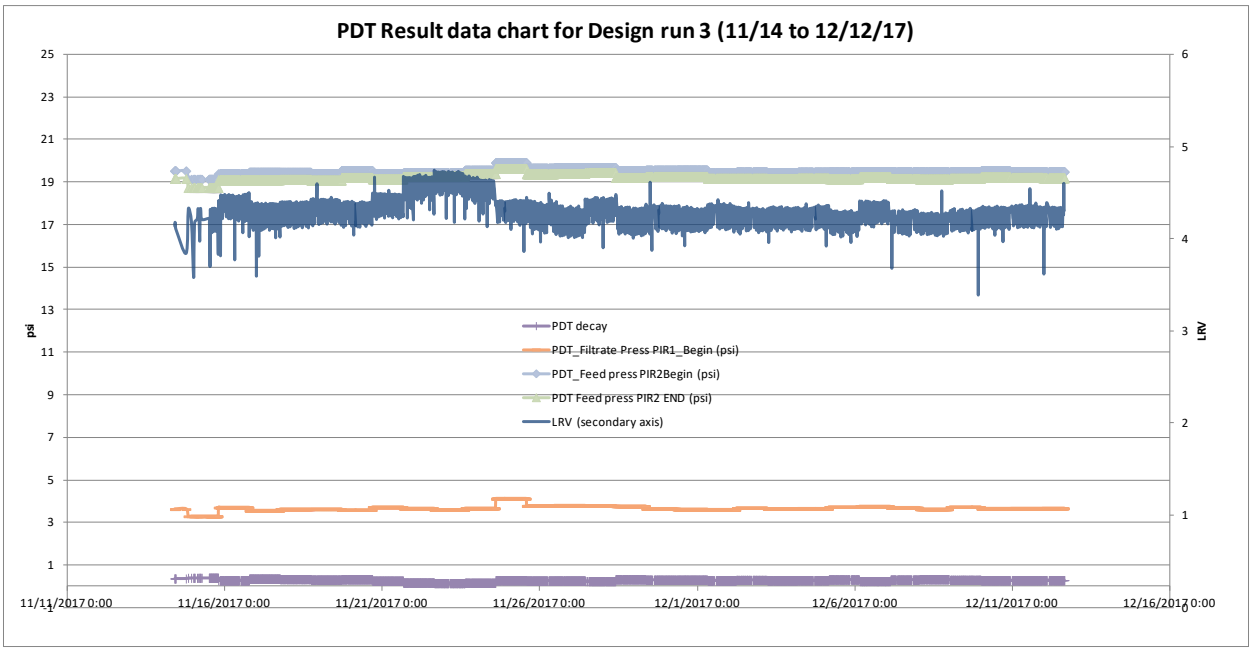


Chart 5.4.1f: Feed Temperature and pH Data for Design Run 3



Note: A PDT result of <1.0 psi drop over 10 minutes assures a 4-log removal of 3 micron sized particles.

Chart 5.4.1g: PDT Result Data for Design Run 3

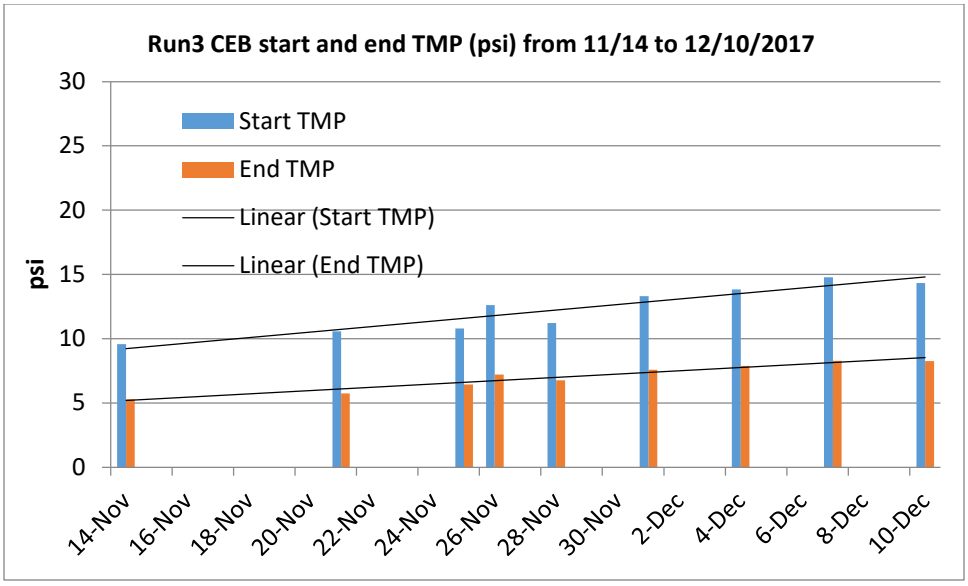


Chart 5.4.1h: CEB Results Data for Design Run 3

6. Discussion

AASI conducted three design run pilot tests at the City of Ashland, OR WTP from August to December 2017. Design run 1 was conducted on TID water source, whereas design runs 2 and 3 were conducted using water from the Reeder Reservoir. Feed water for part of design run 2 and all of design run 3 was pretreated with 8 ppm of coagulant (as 50% ACH), which was injected upstream of all three pilots. Based on data where coagulant feed was consistently dosed, the 8 ppm ACH permitted higher flux operation and with controlled TMP. Lower doses can/should be evaluated.

Design run 1 was a challenging run, with relatively high TOC and turbidity levels found in the TID water source. Feed TOC ranged from 2.4 to 2.9, and feed turbidity ranged from 5 to 20 NTU. By contrast, design run 2 - using the Reeder Reservoir as feed source - had about half the feed TOC levels (1.4 to 1.9 mg/l) and a fraction of the feed turbidity levels (ranging from 0.2 to 2 NTU). During design run 3, however, the Reeder Reservoir TOC level was nearly identical to that of the TID source (ranging from 2.2 to 2.9 mg/l), and the turbidity more than doubled that seen during run 2 (ranging from 0.6 to 3 NTU).

Baseline TMP (i.e., post-backwash TMP) was maintained (average of 3 psi) throughout test run 1, providing a near zero TMP average rate increase / day. However, TMP peaks between CEB events spiked as high as 25 – 30 psi (between 9/14 – 9/16/2017). This may be due to higher feed TOC levels during this period with no coagulant addition. To mitigate more frequent TMP spiking, a dose of 2 ppm of coagulant (as 50% ACH) was injected into the CM3 unit's mix tank on the evening of 9/16/2017. As seen on the TMP chart for design run 1, TMP became more stable soon after the 2 ppm ACH dose started. In general, water with high TOC requires well-controlled pretreatment for stable membrane operation. When operated without coagulant or with insufficient coagulant dosage, the results show faster membrane clogging will occur, requiring a more frequent CIP or more aggressive CEBs.

This run experienced some mechanical and PLC program issues. At the start, we lost data from 8/30 to 9/03/2017 due to data collection rollover. The data cache was full and, to make space for new data, erased previous data in the process. Next, the systems' PDT was not functioning properly and required a PLC program change as well as repair of air leaks around mechanical seals. Also, CEB functionality was impaired at a couple points throughout the test (noted as point C on the TMP chart for design run 1). This problem was found to be a priming issue with the hypochlorite dosing pump. Last, for more than half of the test, filtrate turbidity was experiencing spiking beyond the expected maximum of 0.1 NTU. Turbidity spikes higher than 5 NTU were recorded until it was discovered that the turbidimeter unit was impacted by vibrations from the air compressor pump, which was mounted on the skid along with the turbidimeter. To correct the problem, the turbidimeter unit was relocated off-skid where it was isolated from any vibrations. This modification was made on 9/14/2017 and the resultant stable filtrate turbidity is shown on the filtrate turbidity chart, averaging around 0.015 NTU.

This run shows that despite some problems along the way, and with inconsistent pretreatment, baseline TMP was moderately stable at 50 gfd. The data suggests there is great potential for a higher flux with a controlled and optimally dosed coagulant – the 8 ppm of coagulant (as 50% ACH) injected by the Owner during the last part of run 2 and during run 3 may be higher than necessary. For run 2, when

8 ppm ACH was dosed, we saw a nearly flat TMP response at 50 gfd flux (i.e., from 10/11/17 to 10/31/17). For this flux, coagulant dose could be reduced, or alternatively, a higher flux with the 8 ppm ACH could be achieved, as shown in run 3 TMP data.

Design run 2 was planned as a moderate flux run using the Reeder Reservoir water source. For this run, we decided on a conservative flux of 50 gfd, since coagulant injection was inconsistently applied during this run. Later into the test run, when the coagulant injection was started (on 10/11/2017); we found the flux of 50 gfd to be too conservative and not challenging to the system. This can be seen on the TMP chart from 10/13 to 10/23/2017. To maximize the CM3 performance during this run, we increased the recovery by increasing the filtration / BW cycle time from 60 to 720 minutes. Also, CEB frequency was reduced from 1x/day to 1x/week. The change from 60 to 720 minutes increased percent recovery from 97% to 98.1%.

Starting baseline TMP was at 1.8 psi and ending baseline TMP was 2.19 psi. TMP spiked to 12 psi on 10/06/2017. This event appeared to be due to both no coagulant injection and a weak CEB (i.e., possibly due to pump prime loss). As a result, CEB hypochlorite dosage was increased from 100 to 300 ppm. Later, on 10/11/2017, 8 ppm 50% ACH coagulant was injected into the common feed stream. This change dramatically improved TMP performance and stability, as seen on the TMP data chart for design run 2. CEB frequency was decreased on 10/11/2017 as well – from 1x/day to 1x/week. This reduced the chemical consumption by roughly 30%. Whereas this run has shown TMP to be very stable (i.e., with consistent coagulation), the system was under-challenged. The evidence for this is seen in design run 3, as the system flux was pushed to 150 gfd. A more appropriate “conservative / moderate” flux for design run 2 should have been at least 100 gfd. Filtrate turbidity was recorded between 0 and 0.02 NTU, with average at 0.01 NTU. Permeability appears to have stabilized on 10/13/2017 from 29 to 22 gfd/psi in roughly 18 days.

Design run 3 was planned as the more aggressive flux test. This run also used the Reeder Reservoir water as its feed water source. Coagulant was injected from the start of this test at 8 ppm (as 50% ACH) into the common feed line. Flux was 150

gfd (28 gpm), which was an aggressive flux but with stable TMP. Starting baseline TMP was at 4.9 psi, and ending baseline TMP was 8.3 psi. TMP average rate of increase was 0.15 psi/day. There was a brief period from 11/25 to 11/27/2017 where the coagulant injection was stopped, this caused TMP spiking during this period as shown on the TMP chart. Later on 11/27/2017, coagulant injection was restarted and stable TMP quickly recovered. Permeability at the start of test was 39 gfd/psi and ended at 18 gfd/psi. Based on these results, a 30 day CIP would be recommended, but possibly could go to 60 days (more data would be necessary for more accuracy). Filtrate turbidity was measured below 0.02 NTU throughout the test, except for a brief period of spiking to 0.09 NTU between 1/21 and 1/23/2017. This was a result of the turbidimeter being bumped while doing work nearby. Thereafter, filtrate turbidity stabilized below 0.02 NTU with an average around 0.01 NTU. The PDT chart shows that all PDTs passed. Average decay rate was 0.28 psi /10 mins (i.e., well below the 0.99 psi threshold, where it would fail PDT (Refer to Appendix F). The maximum PDT decay was 0.39 psi, and the minimum PDT decay was 0.14 psi.

7. Conclusions

In conclusion, the series of pilot tests revealed the following optimal operating conditions for the AASI ceramic membrane system under the given conditions and with two different feed water sources:

1) For TID water source

- 50 gfd flux (Note: based on the data, we believe a higher flux can be achieved on this water with consistent and optimal coagulant dosing)
- CEB @ 1x/day, alternating between a) 1-2 pH for 20 minutes followed by 300 ppm NaOCl for 40 minutes, and b) 300 ppm NaOCl for 60 minutes
- 45 minute Backwash / Filtration cycle
- 30 day CIP (or longer – based on when 15 psi baseline TMP is reached, which wasn't reached during the test)
- For this water, Sodium hypochlorite was found to be most effective at cleaning during CIP and CEB

2) For Reeder Reservoir water source

- 50 gfd flux was the conservative flux. This flux performed very well for this water. Based on the data, we believe a higher conservative flux can perform very well with consistent and optimal coagulant dosing.
- 150 gfd flux was the aggressive flux. This flux performed very well for this water, even with a significantly higher feed TOC and turbidity than seen during the lower flux (design run 2). Based on the slow rate of TMP increase, we believe a higher aggressive flux could be used and still have runs longer than 30 days between CIPs.
- CEB @ 1x/3 days when at 150 gfd, whereas 1x/week can be achieved at the conservative flux. CEB was also alternated between a) 1-2 pH for 20 minutes followed by 300 ppm NaOCl for 40 minutes, and b) 300 ppm NaOCl for 60 minutes.
- 720 minute Backwash / Filtration cycle for the conservative flux; whereas 90 minutes Backwash / Filtration cycle for the aggressive flux was ideal.
- CIP cycle for conservative flux can, when extrapolated out, be up to 6 months. CIP cycle for the aggressive flux can be up to 2 months. This is an estimate as this wasn't the focus of the testing. To provide a more accurate frequency would require a longer test period.

8. 30-Year Life Cycle Cost

As noted in the Section 2: Summary, we recommend the full-scale system consist of (2) 12-row units with (10) modules per row, which will provide a total of 6,000 m^2 (64,590 ft^2) of membrane area. A summary of the system's equipment is given in Appendix H.

Based on this approach, the 30-year life cycle costs are shown in Table 8-1. Power and chemical calculations that were used to determine these costs are given in Appendices I and J. Because the same membrane has been installed at over 175 drinking water installations beginning in 1998, with only one membrane element replaced during that time, it's likely that the membrane will not require replacement during the 30-year life of the plant; therefore, no membrane replacement cost has been factored in.

Table 8-1: 30-Year Present Worth of 8 MGD Ceramic Membrane System

	Nov-May	Jun-Oct	Total Annual Cost	Cost Over 30 Years	30 Year Present Worth
Total Contract Price					$ 4,122,437
Feed Pump Power Cost[1]	$ -	$ -	$ -	$ -	$ -
Backwash Pump Power Cost[2]	$ -	$ -	$ -	$ -	$ -
Compressed Air Power Cost	$ 4,649	$ 9,086	$ 13,735	$ 412,050	$ 169,759
ACH Cost[3]	$ -	$ -	$ -	$ -	$ -
Hydrochloric Acid Cost	$ 389	$ 556	$ 945	$ 28,350	$ 11,680
Sodium Hypochlorite Cost	$ 118	$ 162	$ 280	$ 8,400	$ 3,461
Citric Acid Cost	$ 796	$ 758	$ 1,554	$ 46,620	$ 19,207
Chemical Waste Disposal Cost[4]	$ -	$ -	$ -	$ -	$ -
Backwash Waste Disposal Cost[4]	$ -	$ -	$ -	$ -	$ -
Membrane Replacement[5]	$ -	$ -	$ -	$ -	$ -
	TOTAL 30 YEAR PRESENT WORTH =				**$ 4,326,543**

[1] No feed pumps are required as the water will flow by gravity to the membranes.
[2] No backwash pumps are required as the backwash water is propelled pneumatically.
[3] No supplemental coagulant will be needed.
[4] No basis of cost was given in the test protocol.
[5] The ceramic membrane elements should not need replacement.

The present worth of the power and chemical costs (PW) was calculated using the actual costs incurred over 30 years (AC), an annual interest rate of 3%, and the 30-year term using the following equation:

$$PW = AC * (1 / [(1 + 0.03)^{30}])$$

Besides membrane longevity, the other major advantage of our membrane is its high recovery, resulting in a very low cost for handling the chemical and backwash wastewater. Unfortunately, however, the test protocol didn't incorporate this cost in its present worth criteria. If the City wishes to determine this cost, it's estimated that the full-scale system will generate 40,512 gallons per year of chemical wastewater and 22,739 gallons per year of backwash wastewater.

APPENDIX A: Clean-in-Place (CIP) Procedure

As operation proceeds, foulant material will accumulate on the membrane surface which results in low permeability. To recover permeability, a Clean-in-Place (CIP) procedure is required. The following steps detail the procedure.

- **Clean-in-Place (CIP) Procedure**

The objective of a CIP is to restore the baseline permeability of the membrane.

The CIP procedure requires a 25 gallon tank, pump and tubing. The chemicals required are: 1% citric acid, 1000-3000 ppm sodium hypochlorite, (sodium bisulfite and sodium hydroxide, for neutralization, if required).

The CIP process for pilot systems is prepared and operated manually, and includes two separate steps: (**step 1**) 1% citric acid CIP followed by **(step 2)** a 3000 ppm sodium hypochlorite CIP. For both steps, the chemical solution is mixed to 94 L of heated (80 to 100 deg. F, if necessary for deeper cleaning) filtrate in a tank.

To make the 1% citric acid solution, measured 1.88L of 50% liquid citric acid and mix it to 94 liters of (heated, in cold temperatures) filtrate water. To make the 3000 ppm sodium hypochlorite solution, measure 2.25 liters of 12.5% sodium hypochlorite stock solution and mix it to 94 liters of (heated, in cold temperatures) filtrate water.

A pump then transfers the solution into the membrane module feed side which is recirculated through to the top of the module and returns to the tank. Recirculation occurs for 2 hours in cross-flow mode. After 2 hours, the pump is stopped and a valve is closed on top of the module to allow for a 2 hour static soak for both the acid and hypochlorite, separately.

Between each CIP step, a rinse cycle occurs for 10-30 minutes to ensure no chemical mixture occurs. At the end of the (step 2) hypochlorite CIP step, system operation proceeds to filtration mode where TMP can be observed / recorded and permeability is calculated. At the end of each CIP step, drain the module contents to the sewer.

The procedure:

1. Prepare the step 1 CIP acid solution to make 94 liters total volume. Use the tank provided.

2. Set up pump and tubing. Connect tubing to bottom drain of CIP tank and opposite end to pump. Use smaller diameter tubing (with two small valves) in "tee" flow path, to connect to pump output and feed inlet of filter module. Refer to Figure A-1 for flow path to feed the module.

3. Stop process operation.

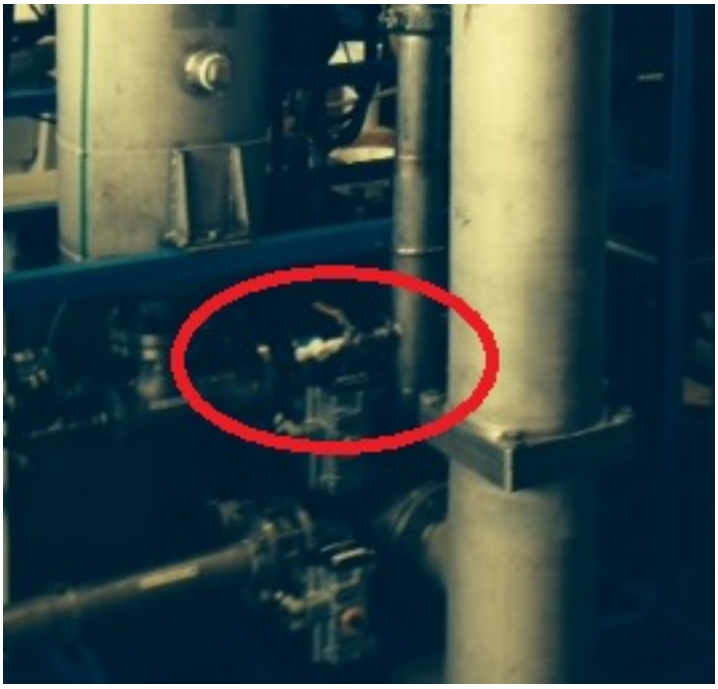
4. **Drain BW tank:** open AV3, AV7 and AV12 (low pressure air) to pressurize and push all fluids to drain. Then close the same valves in reverse sequence (i.e., AV12, AV7, and AV3) when all water has evacuated from the BW tank. For valve locations, refer to the pilot's Process and Instrumentation Drawing, #PID-001.

5. **Drain Membrane module:** open AV3, AV4, AV6 and AV12.

6. Knock on the module housing to know when the contents have drained (i.e., hear a hollow sound).

7. Close all valves. Press "All Auto" button on lower right side of HMI screen in "Operation" page

8. **Release air pressure from module:** Open ball valve "V34" on filtrate side of the module. This valve is located on the left side of the module, labeled as filtrate sample port (about waist high). There should be black tubing connected to it. Place the tubing into a bucket and slowly open the valve by hand. The air should be released. Be careful.



9. Close valve. The module is now ready to be filled.

10. Open valve "V15" on top of module. This is a small, ½" or 3/8" valve with clear tubing connected to it. Place the end of this tube into a bucket.

11. Open the module feed sample port valve, “V20” (on right side of module, just below waist high). You should have the output tubing from the pump connected to here.

12. Open CIP tank valve (located at bottom of CIP tank). Tubing should be connected to here that feeds the suction end of the pump. There are two valves on this tubing. One should be between the pump and the CIP tank, whereas the other is between the pump and to drain.

13. Turn on the pump. The solution should now flow into the module feed side. When you see fluid come out from the top (via the clear tubing), allow some fluid to drain into the bucket (this is to displace any non CIP fluid). Relocate clear tubing (from top of module) to the CIP tank and allow the solution to circulate for 30 minutes to 1 hour (depending on fouling condition).

14. After the circulation period, stop pump, close valves “V15” on top of module (with clear tubing) and “V20” (on bottom feed side of module). This is to allow a static soak for 4-8 for acid cleaning and 8-12 hours for hypochlorite cleaning.

15. At the end of the soak period, drain the module (repeat steps 4-8 above).

16. Prepare “rinse” step: go to “Process” page on the HMI. On the first page, find “Raw Water Feed side”, which should be set to 60 seconds. Change this to 600 seconds.

This is to allow 10 minutes of rinse time to ensure minimal chemical mixing.

17. Go back to "Run/Stop" screen on HMI and press "On" button to start filtration process. This will now fill the feed side for 10 minutes. Afterwards, press "Off" to stop the process.

18. Drain module again, repeat steps 4-8. Remember to release air pressure on the filtrate side valve (step 8).

19. The module is now ready for the second CIP step: Hypochlorite

20. Rinse the CIP tank thoroughly with clean water to remove all acid solution. Also be sure to rinse the pump and tubing with clean water as well.

21. Add your hypochlorite solution to 94 L water into the CIP tank

22. Repeat steps 4 – 16

23. After completing CIP and final rinse step, the system is now ready to check for permeability and/or for operator to conduct a clean water flux test.

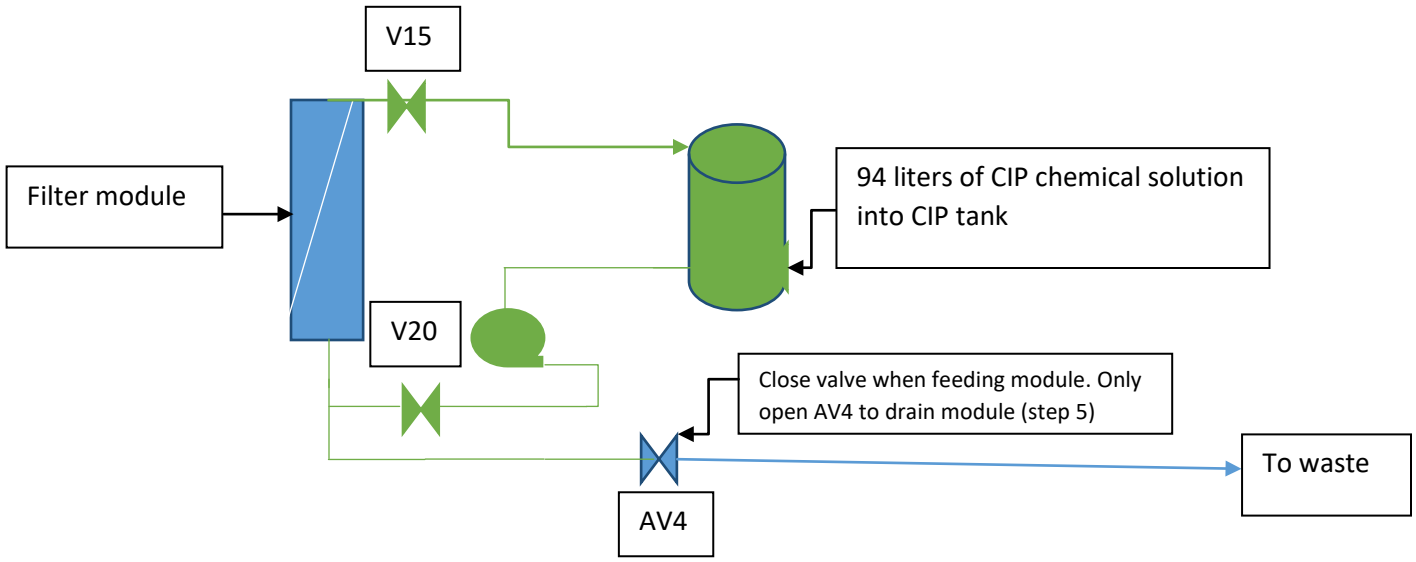


Figure A-1: Recirculation Flow Path

APPENDIX B: Chemically-Enhanced Backwash (CEB) Procedure

The CEB process for the CM-3 pilot system was operated automatically through HMI settings. For this pilot test, the CEB was conducted with various frequencies depending on the Design run. CEB frequency ranged from 1x/day, 1x/3 days, to 1x/ week for both acid (1-2 pH) and hypochlorite (100-300 ppm). In all three design runs, the CEBs alternated between a) a 20-minute soak at 1-2 pH followed by a 40-minute hypochlorite soak, and b) a 60-minute hypochlorite soak.

For the acid CEB, diluted hydrochloric (muriatic) acid was pumped into the backwash tank to make a 1-2 pH solution with filtrate. The CEB process then occurs automatically per the settings from the HMI panel. After the BW tank is filled with filtrate and acid solution, it is then pressurized to 20 psi. When the module is filled with the CEB solution, the pH is measured to ensure the correct dosage was achieved. After the 20-minute soak time, the system proceeds to normal backwash mode (i.e., pressurized to 72.5 psi). After finishing a backwash the system proceeds to the hypochlorite CEB (see below).

For the hypochlorite CEB, diluted sodium hypochlorite was pumped into the backwash tank to make a 100-300 ppm concentration with filtrate. The CEB process then occurs automatically per the settings from the HMI panel. After the BW tank is filled with filtrate and sodium hypochlorite solution, it is then pressurized to 20 psi. When the module is filled with the CEB solution, we measured pH and free chlorine to ensure the correct dosages were achieved. After 40 minutes soak time, the system proceeds to normal backwash mode (i.e., pressurized to 72.5 psi). After finishing a backwash the system proceeds to filtration mode.

The waste from the CEB process is discharged to the BW waste collection tank (on the skid), where it can be neutralized or simply drained out (as it was done during the pilot test at Ashland).

APPENDIX C: Integrity Test Procedure

Membrane integrity was monitored daily to ensure an integral barrier is consistently provided by the membrane surface. Pressure Decay Testing (PDT) is applied as the direct integrity testing method. The process: After backwashing, compressed air enters the filtrate side of the membrane module. After all water is purged by pressurized air, pressure drop is monitored over 10 minutes from an applied air pressure of 20 psi. Results less than 1 psi drop over 10 minutes assures a 4-log removal of 3 micron sized particles.

Procedure:

1) After backwashing, Stop system
2) Adjust AR2 regulator valve to 20 psi
3) Pressurize the filtrate side by opening AV11, AV3, and AV4
4) Ensure that no more water is evacuating from AV4
5) Close AV4, AV3 and AV11
6) Record start pressure, PT2 (filtrate pressure)
7) Set timer to 10 minutes
8) Record end pressure after 10 minutes has expired, PT2 (filtrate pressure)

APPENDIX D: Certifications

- NSF/ANSI 61 certified: for 1 element and 12 element modules. 1 element to minimum flow of 10,000 L /day. 12 modules to min of 120,000 L/day
- California Department of Public Health (CDPH)
  Flux 175gfd, LRV4 *Cryptosporidium*, LRV1 Virus
- CDPH Title 22 for wastewater treatment
- Colorado Department of Public Health and Environment

APPENDIX E: Instrumentation Calibrations

1. Turbidimeters
   a. Feed (Hach ultraturb): via Hach SC1000 controller
      i. Hach StablCal method, 20NTU standard solution
      ii. Time and date of calibration: 08/24/2017 @ 09:31
      iii. Value: 20 NTU
      iv. Gain: 1.61

   b. Filtrate (Hach Ultraturb): via Hach SC1000 controller
      i. Hach StablCal method, 809.0 +/- 10.0 mNTU Standard solution (lot#A6242, exp. Aug. 2018)
      ii. Time and date of Calibration: 08/24/2017 @ 10:11:26
      iii. Value: 0.809mNTU
      iv. Gain: 0.98

2. pH probe: via Hach SC1000 controller
   a. 2 point Calibration to pH7 and pH 10
   b. Time and date of calibration: 08/30/2017 @ 7am

APPENDIX F: PDT results and LRV calculations

Membrane integrity was monitored to ensure an integral barrier was consistently provided by the membrane surface. Pressure Decay Testing (PDT) was applied as the direct integrity testing method.

The process: After backwashing, compressed air enters the filtrate side of the membrane module. After all water is purged by pressurized air, the pressure drop is monitored over 10 minutes from an applied air pressure of 20 psi. Results less than 1 psi drop over 10 minutes assures a 4-log removal of 3 micron sized particles.

Pressure proof and airtight inspection conditions (per manufacturing facility, NGK)

a. Sub-module (without membrane)
    i. Water to 0.75 MPa (108.77psi) / 30 minutes
b. Housing (with membrane)
    i. Air to 0.12 MPa (17.4psi) / 1 min
c. Sub-module (with membrane)
    i. Water to 0.5 MPa (72.5psi) / 30 minutes

LRV Challenge Test

Testing was conducted by MWH to establish and confirm the 4-log *Cryptosporidium* removal ability of the AASI / MWR (NGK) ceramic membrane (*all virus challenge tests were conducted without pretreatment using coagulant in order to conservatively evaluate the virus rejection ability of the ceramic membrane, alone*) by the calculation method, as specified in the Long-Term 2 Enhanced Surface Water Treatment Rule (LT2ESWTR) and Environmental Protection Agency (EPA) Membrane Guidance Manual. Results successfully demonstrated that the AASI / MWR (NGK) ceramic membrane could reliably achieve at least 4-log removal for *Cryptosporidium* (without coagulant) based on the DIT.

LRV Test objectives:

a) *Giardia* Removal: Demonstrate a minimum of 2-log removal of *Giardia* sized particles (e.g., 5-15 µm)
b) *Cryptosporidium* Removal: Demonstrate removal of *Cryptosporidium* sized particles (e.g., 2-5 µm).
c) Virus Removal: Demonstrate a minimum 1-log removal
d) Operations Reliability: Develop data for demonstrating the reliability of various membrane filtration technologies
e) Quality Control: Document a procedure for quality assurance and quality control sufficient to ensure the integrity of the data collection and process.

Challenge test results and calculations

Testing Approach

NGK pressure-driven ceramic membrane pilot unit is operated in a deposition (i.e., dead-end) mode hydraulic configuration. A pressure decay test with an initial pressure of 20 psi was applied to the pilot unit. The operating parameters are listed in Table 5.1 of the MWH report (2005) and detailed in the "Sample Calculation" section below. With the DIT parameters established, the removal of *Cryptosporidium* was calculated and compared using the following models:

- **Hagan-Poiseuille model:** Flat sheet / hollow fiber module, laminar flow through the breach.
- **Orifice model:** Flat sheet module, turbulent flow through the breach.

Main skid calculations (these were test conditions used to determine LRV as conducted by third party testing organization, MWH)

Assumptions:

Operational parameters

- The design capacity of the pilot membrane unit is 36.6 gpm.
- The design flux of the pilot membrane unit is 200 gfd.
- The maximum anticipated water temperature is 77 ^{o}F (25 ^{o}C).
- The minimum anticipated water temperature is 41 ^{o}F (5 ^{o}C).
- The maximum anticipated backpressure (BP_{max}) that might be exerted on the units during direct integrity testing is 78 inches of water.
- The backpressure measured prior to the most recent pressure decay test was 78 inches of water.
- The filtrate flow measured prior to the most recent pressure decay test was 36.6 gpm.

Direct integrity test parameters

- The volume of pressurized piping during the test is 22 L.
- The initial applied test pressure of most recent pressure decay test is 20 psi.
- The duration of the pressure decay test is 10 minutes.
- Baseline (i.e., diffusive) decay is negligible over the duration of the test.

• The smallest verifiable rate of pressure decay under known compromised conditions is 0.10 psi/min.

• The most recent pressure decay test yielded a result of less than 0.10 psi/min.

• The temperature of both the water and the applied air were 74 ^{o}F (23 ^{o}C) during the most recent pressure decay tests.

Unit and membrane characteristics

• The maximum operating TMP is 55 psi.

• The pore shape correction factor (κ) for the membrane material was not determined experimentally, and thus a conservative value of 1 is assumed.

• The liquid-membrane contact angle (θ) was not determined experimentally, and thus a conservative value of 0 is assumed.

Calculate:

1. The minimum direct integrity test pressure commensurate with the required resolution of 3 µm for the removal of *Cryptosporidium*

2. The sensitivity of the direct integrity test

3. The UCL (Upper Control Limit) for the system to achieve a minimum removal of 4 log

4. The LRV verified by the most recent direct integrity test

Solution:

1. Calculate the Minimum Direct Integrity Test Pressure (P_{test}) Commensurate with the Required Resolution of 3 µm for the Removal of *Cryptosporidium*

P_{test} = (0.193 • κ • σ • cosθ) + BP_{max} Equation 4.1

κ = 1 most conservative value (from given information)

σ = surface tension at the air-liquid interface = 74.9 dynes/cm (the surface tension of water at 5 ^{o}C)

θ = 0° most conservative value from given information)

BP_{max} = 78 inches (of water column) from given information

$$P_{test}=\left[\left(0.193psi\cdot\frac{cm}{dyne}\right)*1.0\cdot 74.9\left(\frac{dynes}{cm}\right)\cdot\cos(0°)\right]+\left[\frac{78inches\cdot H2O}{27.7inches\cdot H2O}/psi\right]$$

$\therefore P_{test}$=17.3psi

The pressure calculated above is the lowest permissible initial pressure to apply for the proper test resolution, assuming the minimum anticipated temperature is 5°C. **Table 7-1** lists the P_{test} for required resolution at different minimum anticipated temperatures.

Table F-1: Required Testing Pressure for Proper Resolution at Different Temperatures

Minimum anticipated temperature (°C)	P_{test} (psi)
2	17.4
5	17.3
10	17.1
15	17.0

In the most recent integrity test, since the applied test pressure was 20 psi, the resolution requirement of the LT2ESWTR is satisfied.

2. Calculate the Sensitivity of the Direct Integrity Test

$$LRV_{DIT}=log\left[\frac{Qp\cdot ALCR\cdot Patm}{\Delta Ptest\cdot Vsys\cdot VCF}\right] \qquad \text{Equation 4.9}$$

Q_p = permeate flow = 36.6 gpm (design capacity filtrate flow per membrane)

P_{atm} = atmospheric pressure = 14.7 psi (atmospheric pressure at sea level)

ΔP_{test} = change in test pressure = 0.10 psi/min (smallest verifiable decay rate)

V_{sys} = system volume = 22 L (from given information)

VCF = volumetric concentration factor = 1 (standard for deposition mode {i.e. dead-end filtration] configuration)

ALCR = air-liquid conversion ratio = To be determined

First, Hagan-Poiseuille flow model is used to calculate ALCR. This model is commonly applied to laminar flow breach condition in flat sheet or hollow fiber membrane filtration systems.

ALCR=$\frac{527\cdot\Delta Peff\cdot(175-2.71\cdot T+0.0137\cdot T^2}{TMP\cdot(460+T)}$

Equation C.15

TMP = 55 psi maximum TMP

T = 77 ^{o}F maximum temperature

ΔP_{eff} = effective integrity test pressure = To be determined

$$\Delta Peff = [(Ptest - BP)] * \left[\frac{(Ptest + Patm) + (BP + Patm)}{2 \cdot (BP + Patm)}\right] * \left[\frac{(BP + Patm)}{Patm}\right]$$

Equation C.12

P_{test} = 20 psi initial test pressure

P_{atm} = 14.7 psi atmospheric pressure at sea level

BP = 78 inches (of water column) from given information

Substituting values into Equation C.12 yields ΔP_{eff} = 30.5 psi

The ALCR can be calculated follows:

$$ALCR = \frac{527 \cdot 30.5psi \cdot (175 - 2.71 \cdot 0.0137 \cdot 77^2)}{55psi \cdot (460°\text{F} + 77°\text{F})}$$

∴ALCR =25.9

Substituting values into Equation 4.9 for sensitivity:

$$LRV_{DIT} = \log\left[\frac{(36.6gpm \cdot 3.785\frac{L}{gal}) \cdot 25.9 \cdot (14.7psi)}{0.1\frac{psi}{min} \cdot 22L \cdot 1}\right]$$

$\therefore LRV_{DIT} = 4.4$

Moreover, if a breach is specified as turbulence flow, Orifice flow model is used for the flat sheet system. This generates an even more conservative value of ALCR, which is shown as follows:

Equation C.9

$$ALCR = 170 \cdot Y \cdot \sqrt{\frac{(P_{test} - BP) \cdot (P_{test} + P_{atm})}{(460 + T) \cdot TMP}}$$

TMP = 55 psi maximum TMP

T = 77 ^{o}F maximum temperature

P_{test} = 20 psi initial test pressure for Run-1

P_{atm} = 14.7 psi atmospheric pressure

BP = 78 inches (of water column) from given information

Y = net expansion factor = To be determined

Equation C.10

$$Y = 1 - \left[0.293 * (1 - \frac{BP + P_{atm}}{P_{test} + P_{atm}})\right]$$

Substituting values into Equation C.10 yields Y = 0.85

The ALCR can be calculated follows:

$$ALCR = 170 \cdot 0.85 \cdot \sqrt{\frac{(20psi - \frac{78inches \cdot H2O}{27.7 \frac{inches \cdot H2O}{psi}}) \cdot (20psi + 14.7psi)}{(460°\mathrm{F} + 77°\mathrm{F}) \cdot 55psi}}$$

$$\therefore ALCR = 20.6$$

3. Substituting values into Equation 4.9 for sensitivity:

$$LRV_{DIT} = \log\left[\frac{(36.6gpm \cdot 3.785\frac{L}{gal}) \cdot 20.6 \cdot (14.7psi)}{0.1\frac{psi}{min} \cdot 22L \cdot 1}\right]$$

$$\therefore LRV_{DIT} = 4.3$$

Therefore, at an initial test pressure of 20 psi, the maximum removal value that this membrane filtration system is capable of verifying is 4.4 log using Hagan-Poiseuille model and 4.3 log using Orifice model.

4. Calculate the Upper Control Limit (UCL) For This System

$$UCL = \frac{Q_p \cdot ALCR \cdot P_{atm}}{10^{LRC} \cdot V_{sys} \cdot VCF}$$

Equation 4.17

Q_p = 36.6 gpm design capacity filtrate flow

ALCR = 25.9 using Hagan-Poiseuille mode

ALCR = 20.6 using Orifice model as determined in part 2, above

P_{atm} = 14.7 psi atmospheric pressure at sea level

V_{sys} = 22 L from given information

LRC = log removal credit = 4 to achieve a minimum removal value of 4 log.

VCF = 1 standard for deposition mode (i.e. dead-end filtration) configuration

$$UCL_{Hagan-Pois} = \frac{(36.6gpm \cdot 3.785\frac{L}{gal}) \cdot 25.9 \cdot 14.7psi}{10^4 \cdot 22L \cdot 1 \cdot} = 0.26psi/min$$

$$UCL_{orifice} = \frac{(36.6gpm \cdot 3.785\frac{L}{gal}) \cdot 20.6 \cdot 14.7psi}{10^4 \cdot 22L \cdot 1 \cdot} = 0.20psi/min$$

Therefore, to demonstrate a LRV greater than 4, the pressure decay rate from the integrity test should be less than 0.26 psi/min according to the Hagan-Poiseuille model, and less than 0.20 psi/min according to the Orifice model.

5. Calculate The LRV Verified By The Most Recent Integrity Test.

Equation 4.9 can be used to determine the LRV verified by the most recent integrity test by applying values for the variables specific to the test events.

For the most recent integrity test at an initial testing pressure of 20 psi, the measured test decay rate is less than the smallest verifiable decay rate (0.10 psi/min). Therefore, the membrane unit demonstrates a removal value of 4.4 log using Hagan-Poiseuille model and 4.3 log using Orifice model.

Integrity Test Conclusions

The following conclusions can be made from the testing:

Table F-2 summarizes results from the direct integrity test. NGK established 20 psi as the standard applied pressure for pressure decay DITs. The high mechanic strength of ceramic membrane allows the use of higher pressure (20 psi) with the pressure decay testing, providing the benefit of demonstrating higher LRV.

Table F-2: Summary of Testing Result for main skid

P_{test} (psi)	Model	Module and Defect Flow	Sensitivity of DIT (LRV_{max})	Verified LRV by DIT
20	Hagan-Poiseuille	Flat sheet/ hollow fiber laminar	4.4	4.4
	Orifice	Flat sheet/ turbulent	4.3	4.3

It is important to note that Table F-2 presents conservative values of LRV since the most conservative membrane parameters are used in the calculation. For example, the maximum rated TMP of 55 psi (Colorado Department of Human Services testing condition) was used although the manufacturer

typically suggests operating the ceramic membrane up to 29 psi, under normal operating conditions. As 55 psi generates a lower LRV (4.4-log compared to 4.7-log at 29 psi using the Hagan-Poiseuille model), it is chosen as a more conservative value in the calculation. Moreover, a conservative value is assumed for the pore shape correction factor (κ) and the liquid-membrane contact angle (θ) as they were not determined experimentally.

Table F-3 shows the UCL (Upper Control Limit) values under different conditions to provide a minimum 4-log *Cryptosporidium* removal. Based on the calculation, the allowable pressure decay rate should be less than 0.20 psi/min to demonstrate a minimum removal of 4-log, when a DIT is performed at 20 psi. The manufacturer establishes 0.10 psi/min as their quality control release value (QCRV), which is an even more conservative requirement to ensure that LRV4 can be achieved.

Table F-3: Projection of LRV at Different Pressure Decay Rates

P_{test} (psi)	Model	ΔP_{test} (psi/min)	LRV	UCL for LRV greater than 4 (psi/min)
20	Hagan-Poiseuille	0.1	4.4	0.26
		0.2	4.1	
		0.3	3.9	
	Orifice	0.1	4.3	0.2
		0.2	4.0	
		0.3	3.8	

APPENDIX G: Chemicals used during testing

1) Aluminum Chlorohydrate (ACH): 50% stock. 2 ppm to CM-3 influent during design run 1 from Sept 17-27, 8 ppm to common raw water during design run 2 from Oct 12-31 and during all of design run 3
2) Hydrochloric (Muriatic) Acid: 32% (20 Bé) stock. To 1-2 pH for CEB.
3) Citric Acid: 50% stock. 1% concentration for CIP
4) Sodium Hypochlorite: 12.5% stock. 100-300 ppm concentration for CEB / 3000 ppm concentration for CIP

Chemical consumption for all three Design runs

A) CEB chemical consumption (28 CEBs for design run 1; 21 CEBs for design run 2; 9 CEBs for design run 3; 58 total CEBs)

   1) Muriatic acid: consumption = 1.6 L / CEB; total = 93 L
   2) 100-300ppm (0.01-0.03%) Sodium Hypochlorite: consumption is 0.12 L / CEB; total = 7 L

B) CIP chemical consumption (4 CIPs were performed throughout the three pilot tests)
   1) 1% Citric Acid = 1.88L / 94L water (per CIP) = 7.52 L total consumed
   2) 3000ppm Sodium Hypochlorite = 2.25 L / 94L water (per CIP) = 9 L total consumed

Note: All Chemicals were provided by the City of Ashland, OR

APPENDIX H: Equipment Summary for Full-Scale System

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Compressed Air System

1 Valve Actuation Compressed Air System(s) will be provided as follows:

- (2) compressor pumps rated for 15 hp each pump with a 240-gallon receiver tank, 2 after coolers, alternating control panel, and wide range pressure switches.
- (1) Coalescing, oil removing filter
- (1) Set of 4 machine vibration isolators
- (1) Adjustable pressure regulator
- (1) Air dryer

Membranes

240 Membrane Modules, each consisting of:

- 25 square meter ceramic membrane modules encased in 304 stainless steel housing.

24 Membrane Module Rack(s), each consisting of:

- Carbon steel structural framework, painted for corrosion protection.
- Piping within the membrane system skid, required for interconnection of skid mounted membrane system components. Piping will be schedule 40 PVC, socket welded or threaded. Air piping will be stainless steel tubing and schedule 10 stainless steel piping.
- Wiring of valves and instruments to skid panel.
- 1/2" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 1" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 2.5 inch pneumatically operated wafer style butterfly valve(s).
- 3 inch pneumatically operated wafer style butterfly valve(s).
- 1.5" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 4 inch pneumatically operated wafer style butterfly valve(s).
- 8 inch pneumatically operated wafer style butterfly valve(s).
- 12 inch pneumatically operated wafer style butterfly valve(s).
- 1/2" stainless steel manual ball valves.
- 1.5" stainless steel check valve.

2 Remote I/O Panel(s) consisting of:

- NEMA 4X fiberglass enclosure(s).
- Fuses and fuse blocks.
- GFI convenience outlet(s).
- Control relay(s).
- Selector switch(es).
- Indicating pilot light(s).
- Ethernet switch(es).
- Power supply(ies).
- Terminal blocks.
- Ethernet adapter module(s).
- Input card(s).
- Output module(s).
- Analog output card(s).
- Terminal base(s).
- UL label(s).
- Disconnect switch(es).

1 Control Panel(s) consisting of:

- NEMA 12 panel enclosure suitable for indoor installation and constructed of painted steel.

- Fuses and fuse blocks.
- GFI convenience outlet(s).
- Control relay(s).
- Selector switch(es).
- Indicating pilot light(s).
- Ethernet switch(es).
- Hubbell Cat 6 Ethernet Cable.
- Power supply(ies).
- Terminal blocks.
- Allen Bradley 1769-L33ER Compactlogix integral programmable controller.
- Compactflash.
- End cap(s).
- Power supply(s).
- Input card(s)
- Relay output(s).
- Analog input(s).
- Analog output(s).
- Panelview plus 6 1250 operator interface(s).
- UL label(s).

1 CIP System(s) will be provided as follows:

- 500-gallon polyethylene tank.
- (2) Centrifugal pump with 1.5 HP, 3 ph. motor.
- 1/2" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 1" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 1 1/2" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 3" manual butterfly valve(s).
- pH/ORP sensor(s).
- Pressure transducer(s).
- Temperature transmitter(s).
- Hach CL17 total chlorine process analyzer(s).

2 Strainer(s) will be provided as follows:

- 18 inch automatic backwashable strainer with NSF-approved epoxy lining and 60 mesh 304 stainless steel screen.

Membrane Accessories

2 Common Skid Valves and Instrumentation, including the folllowing:

- 1/2" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 1" full port, three piece, stainless steel body ball valve(s), grooved end connections with single phase electric actuator(s). Valve / actuator combination shall be manufactured by TCI / Nibco or equal.
- 2.5 inch pneumatically operated wafer style butterfly valve(s).
- 3 inch pneumatically operated wafer style butterfly valve(s).
- 1/4" Threaded bronze ball valve.
- 1/2" stainless steel manual ball valves.
- 3/4" globe valve.
- 1" stainless steel manual ball valves.
- 1.5" stainless steel manual ball valves.
- 2.5" manual butterfly valve(s).
- 3" manual butterfly valve(s).
- 3" Magnetic flow meter(s).
- 8" Magnetic flow meter(s).
- 3/4" check valve.
- 1.5" check valve(s).
- 2.5" check valve(s).
- Pressure control valves.

- 1.5" pressure control valve.
- 1/2" air pressure regulator.
- Pressure transducer(s).
- Air release valve.
- Pressure switches.
- Level transmitter.

2 Backwash System Tank(s) consisting of:

- Painted carbon steel backwash water tank(s).
- Painted steel air receiver tank(s).

Permeate Discharge Components

2 Turbidity Meter Assembly(ies) consisting of:

- Turbidity sensor(s).
- Hach SC200 controller and display module(s).
- Calibration start-up kit(s).
- Turbidity mounting plate(s).

Chemical Feed Systems

3 Chemical Feed Systems for hydrochloric acid, sodium hypochlorite, and citric acid, each system consisting of the following:

- UHMW backing panel(s).
- Calibration columns.
- Chemical feed pumps.
- Pressure relief valves.
- GFI outlet(s).
- Ball valves.
- 316 stainless steel shelf weldments.
- Polypropylene utility trays.

APPENDIX I: Compressed Air Calculations for Full-Scale System

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Compressor Calculations for the 7 Winter Months

Valves	Feed Valves (Top and Bottom)	Filtrate Valves	Backwash Inlet Valves	Drain Valves (Top and Bottom)	Row Isolation Valves	Backwash Pump Discharge Valves
Cycles per hour	0.28	0.28	0.28	0.28	0.28	0.28
Total Qty	4	26	96	0	48	48
Valve Selected (same size as pipe)	2.5	3	4	6	8	12
(N=1.25 above catalog values)	**Double Acting Actuators/ Reduced Disc Dia Valves**		**(Fail in Position)**			
Actuator (N=1.25 above catalog torque req'd)	92-0630	92-0830	92-0830	92-0830	92-1180	92-1270
Basic Size Number	63	83	83	83	118	127
Air req'd for full forward stroke (in^3)	9.6	24.8	24.8	24.8	73.8	96.7
Air req'd for full return stroke (in^3)	13.4	32.6	32.6	32.6	95.5	130.8
Hourly Air Requirement (in^3)	26	416	1538	0	2268	3047
Total Air Requirement (including 10% for leakage, acfm)			0.08			
Total Std Air Requirement (including 10% for leakage, scfm) @ 80 psig			**0.55**			
			default values			
Compressed Air Pressure (High)	90	psig	90			

TANK VOLUME			
Rest Time =	30	Min	30
Cut out PSI standard Switch =	110	PSI	110
Cut out PSI high capacity switch =	130	PSI	130
Cut in PSI =	80	PSI	80

105 psi average in receiver

Valve Compressor Calculations

14.2097386	psia compressor inlet	
0.59	icfm compressor inlet total	
70%	compressor efficiency	
0.15900983	bhp total	
2.84576719	kwh/day total	

Backwash Compressor Calculations

32	L/m2 @ 14.7 psia required
17.9166667	min fill time
447	L/min @ 14.7 psia total
15.8	cfm @ 14.7 psia total
14.2	psia compressor inlet
16.3	icfm compressor inlet
70%	compressor efficiency
4.4	bhp per train
156.4	kwh/day total all trains

Cost of Energy (EC) = $0.08 /kwh

Total Compressor Power Cost = $4,648.74 per year

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Compressor Calculations for the 5 Summer Months

Valves	Feed Valves (Top and Bottom)	Filtrate Valves	Backwash Inlet Valves	Drain Valves (Top and Bottom)	Row Isolation Valves	Backwash Pump Discharge Valves
Cycles per hour	0.55	0.55	0.55	0.55	0.55	0.55
Total Qty	4	26	96	0	48	48
Valve Selected (same size as pipe)	2.5	3	4	6	8	12
(N=1.25 above catalog values)	**Double Acting Actuators/ Reduced Disc Dia Valves**		**(Fail in Position)**			
Actuator (N=1.25 above catalog torque req'd)	92-0630	92-0830	92-0830	92-0830	92-1180	92-1270
Basic Size Number	63	83	83	83	118	127
Air req'd for full forward stroke (in^3)	9.6	24.8	24.8	24.8	73.8	96.7
Air req'd for full return stroke (in^3)	13.4	32.6	32.6	32.6	95.5	130.8
Hourly Air Requirement(in^3)	50	814	3006	0	4433	5956
Total Air Requirement (including 10% for leakage, acfm)			0.15			
Total Std Air Requirement (including 10% for leakage, scfm) @ 80 psig			**1.08**			
			default values			
Compressed Air Pressure (High)	90	psig	90			

TANK VOLUME			
Rest Time =	30	Min	30
Cut out PSI standard Switch =	110	PSI	110
Cut out PSI high capacity switch =	130	PSI	130
Cut in PSI =	80	PSI	80

Valve Compressor Calculations

14.2097386	psia compressor inlet
1.16	icfm compressor inlet total
70%	compressor efficiency
0.31079195	bhp total
5.56218132	kwh/day total

Backwash Compressor Calculations

32	L/m2 @ 14.7 psia required
9.16666667	min fill time
873	L/min @ 14.7 psia total
30.8	cfm @ 14.7 psia total
14.2	psia compressor inlet
31.9	icfm compressor inlet
70%	compressor efficiency
8.5	bhp per train
305.6	kwh/day total all trains

105 psi average in receiver

Cost of Energy (EC) = $0.08 /kwh

Total Compressor Power Cost = $9,086.16 per year

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APPENDIX J: Chemical Calculations for Full-Scale System

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CEB Calculations - NaOCl

Project: **Ashland, OR - 7 Winter Months**

A. Fixed Conditions

Number of trains:	2	
Number of modules/train:	120	
Module area:	25	m^2/module
Total Filtration Area/train:	3000	m^2 per train
Chemically Enhanced Backwash (CEB) Flux:	246	L/m^2-hour
CEB Flow rate/train:	12,300	L/min (3249 gpm) per train

B. Chemical Requirements and Assumptions

Targeted NaOCl concentration for CEB3:	300	mg/L
NaOCl dosage:	3690.0	g NaOCl/min per train

C. Dosing Requirements for Selected Chemicals:

Purchased or Targeted Concentration NaOCl:	12.5%	
Weight of NaOCl required:	29,520.0	g of 12.5% NaOCl/min per train
Specific Gravity of NaOCl	1.210	g/mL
12.5% NaOCl dosage Requirement	24397	mL/min (386.7 gal/hour) per train
CEB3		
CEB3 Dosing Time:	1.00	minutes/event
CEB3 Frequency:	0.50	per day per train
CEB3 Consumption:	12.20	liters/day per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	24,397	mL/min (386.7 gal/hour) per train
Chemical Consumption per train:	12.20	liters/day (3.22 gal/day) per train
Total Daily System Consumption:	24.40	liters/day (6.44 gal/day)
Refill time desired:	9	days
Volume of container required:	208	liters (55 gal) or larger
Annual Average Rate of Consumption (7 months):	5,194	L/year (1372 gal/year)
Chlorine Cost:	$0.125	per pound Cl_2
Total annual cost:	$103	per year

CEB Calculations - NaOCl

Project: **Ashland, OR - 5 summer months**

A. Fixed Conditions

Number of trains:	2	
Number of modules/train:	120	
Module area:	25	m^2/module
Total Filtration Area/train:	3000	m^2 per train
Chemically Enhanced Backwash (CEB) Flux:	246	L/m^2-hour
CEB Flow rate/train:	12,300	L/min (3249 gpm) per train

B. Chemical Requirements and Assumptions

Targeted NaOCl concentration for CEB3:	300	mg/L
NaOCl dosage:	3690.0	g NaOCl/min per train

C. Dosing Requirements for Selected Chemicals:

Purchased or Targeted Concentration NaOCl:	12.5%	
Weight of NaOCl required:	29,520.0	g of 12.5% NaOCl/min per train
Specific Gravity of NaOCl	1.210	g/mL
12.5% NaOCl dosage Requirement	24397	mL/min (386.7 gal/hour) per train
CEB3		
CEB3 Dosing Time:	1.00	minutes/event
CEB3 Frequency:	1.00	per day per train
CEB3 Consumption:	24.40	liters/day per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	24,397	mL/min (386.7 gal/hour) per train
Chemical Consumption per train:	24.40	liters/day (6.44 gal/day) per train
Total Daily System Consumption:	48.79	liters/day (12.89 gal/day)
Refill time desired:	26	days
Volume of container required:	1249	liters (330 gal) or larger
Annual Average Rate of Consumption (5 months):	7,421	L/year (1960 gal/year)
Chlorine Cost:	$0.125	per pound Cl_2
Total annual cost:	$147	per year

Project: Ashland, OR - 7 Winter Months

A. Fixed Conditions

Number of trains:	2	
Number of modules/train:	120	
Module area:	25	m^2/module
Total Filtration Area/train:	3000	m^2 per train
Chemically Enhanced Backwash (CEB) Flux:	246	L/m^2-hour
CEB Flow Rate/train:	12,300	L/min (3249 gpm) per train

B. Chemical Requirements and Assumptions

Targeted pH during CEB1/2 (Acid):	2.0		*range is 2 - 2.5 pH*
H^+ Concentration is then:	0.010	M/L	
Molecular Weight of H^+:	1.0	g H^+/M	
Required H^+:	0.010	g H^+/L	
Molecular Weight of HCL:	36.45	g HCl/M	
Theoretical Requirement of HCl:	0.365	g HCl/L	

Note: Preferrably, the amount of HCl (i.e. g/L needed to lower the pH to 2) is performed by titration. In this case, enter the titrated quantity in the following 'Actual Requirement' cell.

Actual Requirement of HCl:	0.365	g HCl/L
The unit rate of HCl addition:	4483	g HCl/minute per train

C. Dosing Requirements for Selected Chemicals:

Purchased or Targeted Concentration HCl:	32%	
Specific Gravity of HCl	20	°Be (Baume')
	1.16	g/mL of 32% HCl
32% HCl dosage requirement:	12,078	mL/min (191.5 gal/hr) per train

CEB2

CEB2 Dosing Time	1.00	minutes/event
CEB2 Frequency:	0.25	per day per train
CEB2 Consumption:	3.02	liters/day per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	12,078	mL/min (191.5 gal/hr) per train
Chemical Consumption per train:	3.0	liters/day (0.8 gal/day) per train
Total Daily System Consumption:	6.0	liters/day (1.6 gal/day) Total
Refill time desired:	34	days
Volume of container required:	208	liters (55 gal) or larger
Annual Average Rate of Consumption (7 months):	1,286	L/year (340 gal/year)
Acid Cost:	$0.37	per pound 100% HCl
Total annual cost:	$389	per year

CEB Calculations - HCl

Project: Ashland, OR - 5 summer months

A. Fixed Conditions

Number of trains:	2	
Number of modules/train:	120	
Module area:	25	m^2/module
Total Filtration Area/train:	3000	m^2 per train
Chemically Enhanced Backwash (CEB) Flux:	246	L/m^2-hour
CEB Flow Rate/train:	12,300	L/min (3249 gpm) per train

B. Chemical Requirements and Assumptions

Targeted pH during CEB1/2 (Acid):	2.0		*range is 2 - 2.5 pH*
H^+ Concentration is then:	0.010	M/L	
Molecular Weight of H^+:	1.0	g H^+/M	
Required H^+:	0.010	g H^+/L	
Molecular Weight of HCL:	36.45	g HCl/M	
Theoretical Requirement of HCl:	0.365	g HCl/L	

Note: Preferrably, the amount of HCl (i.e. g/L needed to lower the pH to 2) is performed by titration. In this case, enter the titrated quantity in the following 'Actual Requirement' cell.

Actual Requirement of HCl:	0.365	g HCl/L
The unit rate of HCl addition:	4483	g HCl/minute per train

C. Dosing Requirements for Selected Chemicals:

Purchased or Targeted Concentration HCl:	32%	
Specific Gravity of HCl	20	°Be (Baume')
	1.16	g/mL of 32% HCl
32% HCl dosage requirement:	12,078	mL/min (191.5 gal/hr) per train

CEB2

CEB2 Dosing Time	1.00	minutes/event
CEB2 Frequency:	0.50	per day per train
CEB2 Consumption:	6.04	liters/day per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	12,078	mL/min (191.5 gal/hr) per train
Chemical Consumption per train:	6.0	liters/day (1.6 gal/day) per train
Total Daily System Consumption:	12.1	liters/day (3.2 gal/day) Total
Refill time desired:	17	days
Volume of container required:	208	liters (55 gal) or larger
Annual Average Rate of Consumption (5 months):	1,837	L/year (485 gal/year)
Acid Cost:	$0.37	per pound 100% HCl
Total annual cost:	$556	per year

CIP Calculations - NaOCl

Project: **Ashland, OR - 7 Winter Months**

A. Fixed Conditions

Number of trains:	2	
Number of modules/train:	120	
Module area:	25	m^2/module
Total Filtration Area/train:	3000	m^2 per train
Volume per Element:	85	L/element
Volume per Train:	11,220	L/train *includes 10% extra for piping*

B. Chemical Requirements and Assumptions

Targeted NaOCl concentration for CIP:	3,000	mg/L
NaOCl dosage:	33660.0	g NaOCl per train

C. Dosing Requirements for Selected Chemicals:

Purchased or Targeted Concentration NaOCl:	12.5%	
Weight of NaOCl required:	269,280.0	g of 12.5% NaOCl per train
Specific Gravity of NaOCl	1.210	g/mL
12.5% NaOCl dosage Requirement	222545	mL/train = (### gallons/train)
CIP		
CIP Makeup Time:	10.00	minutes/event
CIP Frequency:	3.00	per year per train
CIP Consumption:	667.64	liters/year per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	22,255	mL/min per train
Chemical Consumption per train:	667.64	liters/year (176.37 gal/year) per train
Total Yearly System Consumption:	1335.27	liters/year (352.74 gal/year)
Refill time desired:	0.9	years
Volume of container required:	1249	liters (330 gal) or larger
Annual Average Rate of Consumption (7 months):	779	L/year (206 gal/year)
Chlorine Cost:	$0.125	per pound Cl_2
Total annual cost:	$15	per year

CIP Calculations - NaOCl

Project: **Ashland, OR - 5 summer months**

A. Fixed Conditions

Number of trains:	2		
Number of modules/train:	120		
Module area:	25	m^2/module	
Total Filtration Area/train:	3000	m^2 per train	
Volume per Element:	85	L/element	
Volume per Train:	11,220	L/train	*includes 10% extra for piping*

B. Chemical Requirements and Assumptions

Targeted NaOCl concentration for CIP:	3,000	mg/L
NaOCl dosage:	33660.0	g NaOCl per train

C. Dosing Requirements for Selected Chemicals:

Purchased or Targeted Concentration NaOCl:	12.5%	
Weight of NaOCl required:	269,280.0	g of 12.5% NaOCl per train
Specific Gravity of NaOCl	1.210	g/mL
12.5% NaOCl dosage Requirement	222545	mL/train = (### gallons/train)
CIP		
CIP Makeup Time:	10.00	minutes/event
CIP Frequency:	4.00	per year per train
CIP Consumption:	890.18	liters/year per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	22,255	mL/min per train
Chemical Consumption per train:	890.18	liters/year (235.16 gal/year) per train
Total Yearly System Consumption:	1780.36	liters/year (470.32 gal/year)
Refill time desired:	0.7	years
Volume of container required:	1249	liters (330 gal) or larger
Annual Average Rate of Consumption (5 months):	742	L/year (196 gal/year)
Chlorine Cost:	$0.125	per pound Cl_2
Total annual cost:	$15	per year

CIP Calculations - Citric Acid

Project: **Ashland, OR - 7 Winter Months**

A. Fixed Conditions

Number of trains:	2	
Number of modules/train:	120	
Module area:	25	m^2/module
Total Filtration Area/train:	3000	m^2 per train
Volume per Element:	85	L/element
Volume per Train:	11,220	L/train *includes 10% extra for piping*

B. Chemical Requirements and Assumptions

Targeted acid concentration for CIP:	10,000	mg/L
Citric acid dosage:	112200.0	g $C_6H_8O_7$ per train

C. Dosing Requirements for Selected Chemicals:

Purchased Concentration:	50.0%	
Weight of citric acid required:	224,400.0	g of 50% citric acid per train
Specific Gravity of citric acid:	1.250	g/mL
50% citric acid dosage requiremen:t	179520	mL/train = (### gallons/train)
CIP		
CIP Makeup Time:	10.00	minutes/event
CIP Frequency:	3.00	per year per train
CIP Consumption:	538.56	liters/year per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	17,952	mL/min per train
Chemical Consumption per train:	538.56	liters/year (142.27 gal/year) per train
Total Yearly System Consumption:	1077.12	liters/year (284.55 gal/year)
Refill time desired:	0.87	years
Volume of container required:	937	liters (247.5 gal) or larger
Annual Average Rate of Consumption (7 months):	628	L/year (166 gal/year)
Chlorine Cost:	$0.92	per lb 100% citric acid
Total annual cost:	$796	per year

CIP Calculations - Citric Acid

Project: **Ashland, OR - 5 summer months**

A. Fixed Conditions

Number of trains:	2	
Number of modules/train:	120	
Module area:	25	m^2/module
Total Filtration Area/train:	3000	m^2 per train
Volume per Element:	85	L/element
Volume per Train:	11,220	L/train *includes 10% extra for piping*

B. Chemical Requirements and Assumptions

Targeted acid concentration for CIP:	10,000	mg/L
Citric acid dosage:	112200.0	g $C_6H_8O_7$ per train

C. Dosing Requirements for Selected Chemicals:

Purchased Concentration:	50.0%	
Weight of citric acid required:	224,400.0	g of 50% citric acid per train
Specific Gravity of citric acid:	1.250	g/mL
50% citric acid dosage requiremen:t	179520	mL/train = (### gallons/train)
CIP		
CIP Makeup Time:	10.00	minutes/event
CIP Frequency:	4.00	per year per train
CIP Consumption:	718.08	liters/year per train

D. Summary Requirements:

Chemical Feed Pumping Rate:	17,952	mL/min per train
Chemical Consumption per train:	718.08	liters/year (189.7 gal/year) per train
Total Yearly System Consumption:	1436.16	liters/year (379.39 gal/year)
Refill time desired:	0.87	years
Volume of container required:	1249	liters (330 gal) or larger
Annual Average Rate of Consumption (5 months):	598	L/year (158 gal/year)
Chlorine Cost:	$0.92	per lb 100% citric acid
Total annual cost:	$758	per year