Rapport

Diarienummer
NV Rapport 2026-04

Projektnummer
NFVP 4420

Microfiltration (MF)-pilot

Results and experience with ceramic microfiltration membranes for ultrafiltration backwash water treatment

Ramboll
i samarbete med Norrvatten

2026-08-24

Intended for
Norrvatten, NFVP

Type of document
Report

Date
November 2025

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Projektnamn	Norrvatten NFVP Maskin & Process NFVP-013
Projekt nr	1320054148
Mottagare	Kommunalförbundet Norrvatten
Typ av dokument	Rapport
Version	1.0
Datum	2025-11-12
Förberett av	Ida Sekizovic, Julia Wadefjord
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1. Introduction

This pilot project is a part of evaluating whether a microfiltration (MF) technique is relevant to implement as part of Norrvatten's future new water treatment plant (NFVP) as a treatment step for the dirty backwash water from the UF-step. The current proposed second ultrafiltration (UF2) system involves using UF membranes to treat dirty backwash water from the first UF step. When operating at design capacity, this system produces a volume of backwash waste to the existing sludge treatment system that increases the loading (including future extension) to a level in excess of the current loading rate. Moving from UF membrane to submerged MF membranes allows for a system which can handle high levels of suspended solids, produce similar permeate water quality to UF2 process, and increase the recovery thus reduce the amount of waste/sludge that will be pumped to the existing sludge treatment system. Henc, the use of MF ceramic membranes is interesting to investigate in the view of capacity and feasibility of treating dirty backwash water from the UF membranes instead of a second UF step.

In this project, a UF-pilot (delivered from Pentair) filters surface water from Lake Mälaren in combination with direct coagulation. The same method as full-scale. A bench scale MF-pilot was purchased from Split Water Nordic, a company that is a collaborative partner with Cembrane. This pilot is used to filter the dirty backwash water from the UF-pilot. The MF-pilot testing period took place from November 2024 to June 2025.

Figure 1 shows a schematic overview of the MF-pilot process set up.

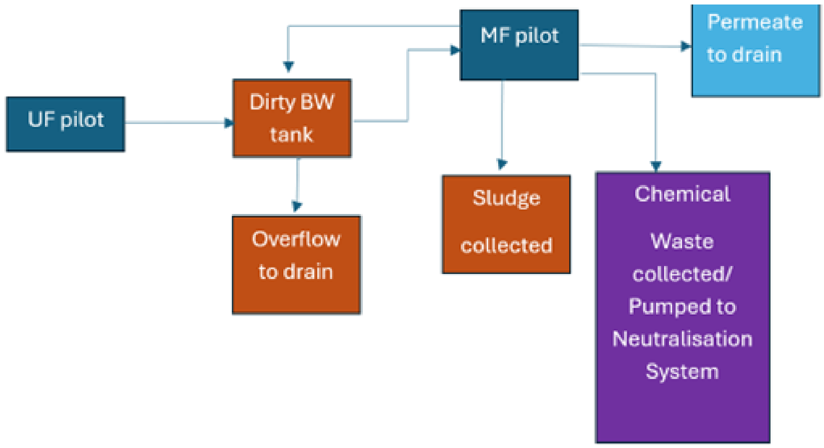

Figure 1. Schematic block diagram overview of the MF-process.

1.1 Purpose and Goals

The purpose of the project is to investigate and evaluate the viability of using ceramic microfiltration membranes to treat the dirty backwash water from the ultrafiltration system. The focus will be on determining an optimal flux, minimizing pressure build up over the membrane, sludge production and the impact of cleaning the membranes with different cleaning solutions. MF-membranes are also examined for its potential to lower the flow to the sludge treatment.

The goals of the project are:

- Identifying a design flux for the treatment of UF1 backwash waste in MF for the full-scale design of NFVP project

- Finding a time interval for when the membranes need chemical cleaning
- Finding a time interval of backpulsing (form of backwashing) of the membranes
- Analyze the permeate and backwash water, both chemical and biological analysis.
- Determine the sludge production and qualities from the process

1.2 Scope

The pilot was operated from November 2024 to June 2025 and initially required intensive manual control. This meant that an operator had to monitor the pilot continuously to manage the start-up procedure, operate valves for degas and backpulse, adjust the pump to maintain a stable flow, and handle the shut-down procedure. Additionally, flow, suction pressure, and time had to be recorded manually. The pilot operated for 4 to 6 hours during a working week.

Minor automation was later installed in the pilot (February 2025), allowing the flow and backpulsing procedure to be automated. This enabled results to be viewed online in real time, with the option to download data and further processing it. However, procedures such as degas, start-up, and shutdown, as well as membrane cleaning, still had to be performed manually. Results will be presented from March 2025 to June 2025, due to stable process operations from UF-pilot and online monitoring of the MF-pilot.

Throughout the trial, the permeate pump experienced issues with air bubbles which impacted the performance negatively, especially when the flow exceeded 100 lmh. Even after a larger permeate pump was installed, no significant improvements were seen at higher flux. This meant that air entered the system when the suction pressure increased, requiring a more frequent degas interval, which was more suited to an automatic degassing. However, limitations in budget and time did not allow for this modification. In a full-scale pilot application or a full-scale plant this problem is addressed with an automatic degassing sequence, so this problem is related only to this small scale mini-pilot.

2. Methodology and Implementation

The methodology and implementation of the project is described in the following chapter.

2.1 Pilot design

The MF-pilot is a bench-scale pilot and has been operated at Norrvatten drinking water treatment plant, Görvälnverket. The supplier of the MF-pilot is Split Water Nordic. In Table 1, technical data from the supplier are presented.

Table 1. Technical data from the supplier.

Number of membranes:	6 ceramic plates, total area 1 m² (lmh = l/h)
Membrane material:	Silicon carbide, SiC
Filtration mode:	Out to in
Capacity:	250 l/h (equivalent to 250 lmh). Maximum suction pressure of -250 mbar
Instruments:	Permeate flow meter, permeate pressure meter

An overview of the pilot, along with a description of its connected components, is presented in Figure 2 and Table 2.

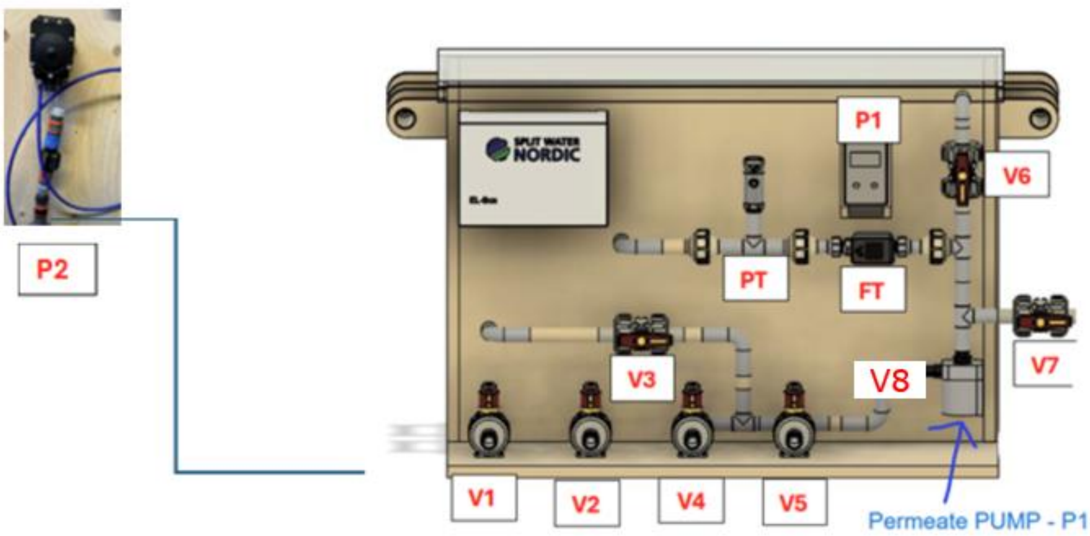

Figure 2. Schematic overview of the pilot and its components.

Table 2. Description of the components in the MF-pilot.

Components	Description
P1	Permeate pump and monitoring display connected to the permeate pump
P2	Membrane pump/feed pump, connected to V1. Sampling point MF FEED
PT	Pressure transmitter, Suction, negative value
FT	Flow transmitter
V1	Infeed Valve
V2	Drain Valve
V3	Not in use (always closed)
V4	Outfeed valve, permeate. Sampling point MF PERM
V5	Not in use (always closed)
V6	Degas Valve
V7	Backwash Valve (drinking water) (magnetic valve)
V8	Flow valve for regulating Permeate Flow

The dirty backwash water from the UF-pilot is collected in a 500 L collector tank, named dirty backwash (BW) tank. Using the feed pump, the dirty backwash water is transferred to the membrane tank, which has an approximate capacity of 150 L. The membrane tank features an overflow outlet that directs excess water back to the dirty BW tank. Filtration is done by the permeate pump (suction) and the system is cleaned by backpulse. Sludge or concentrate can be drained from the membrane filter tank to a floor drain or collected in a receptacle for analysis or further dewatering tests. The permeate water can be directed either to the chemical waste tank or outgoing pipe.

2.2 Operational settings

The membranes used are known as submerged membranes, meaning that they are placed in a tank with the dirty water to be filtered. The inflow of dirty backwash water to the membrane tank is continuous and is adjusted to match the permeate flow rate. The permeate pump, together with the permeate valve, regulates the permeate flow. Degas is performed by closing the permeate valve, turning off the pump, and opening the degas valve for a few seconds.

When performing backpulse to reduce the pressure build-up over the membrane, MF permeate or in this case pressurized drinking water was used. This was first done manually then subsequently automated. A valve after the permeate pump automatically closes, followed by the opening of the solenoid valve from a pressurized drinking water line. This allows drinking water to backflow through the membranes for approximately 30 seconds.

The pilot did not have a continuous sludge drain, the sludge was collected in the bottom of the membrane tank during operation. After each day of operation, the membrane tank was first drained of the clear supernatant over the sludge and then the bottom of tank was emptied of sludge.

During the trial period, various operational settings were tested, focusing on different flux rates, backpulse intervals, and cleaning intervals. The trial period was divided into five phases, each targeting a specific flux rate. Table 3 presents these phases, detailing the targeted flux and required backpulse interval. However, when analyzing the results, the phases will be categorized into cleaning sequences, as elaborated in Chapter 4.

Table 3. Summary of flux and required backpulse intervals for air bubble removal.

Period	Date	Flux [lmh]	Backpulse interval [min]	Description
P1	March 27-31, 2025	80	30	Start-up trial and performance of the membranes
P2	April 1-7, 2025	100	20-30	Problem with the pump to keep the flux at 100 lmh
P3	April 8-9, 2025	150	5-15	It was challenging to maintain a flux of 150 lmh for more than a few minutes
P4	April 13-May 13, 2025	100	20-30	Continue with 100 lmh
P5	May 14-June 12, 2025	70-80	20	Operating 80 lmh Ran overnight during the week; however, during the night, the flux was lowered to 60-70 lmh

2.3 Chemical cleaning procedure or Chemical Enhanced Soak (CES)

The cleaning procedure recommended by the supplier involved the following steps:

1. Spraying the membranes with sodium hypochlorite (NaClO), at a concentration of 1000 mg/L, total volume of 1,40 L. Use permeate water from UF1.
2. Allow a contact time of 1 h
3. Rinse the membranes by filling the pilot tank with clean water using the backwash procedure.
4. Spray the membranes with sulfuric acid (H_2SO_4) adjusted to pH 2, total volume of 1,40 L. Use permeate water from UF1.
5. Allow another 1 h contact time.
6. Perform a final rinse using the backwash procedure.

During the cleaning procedure the membrane tank is emptied, and the membranes are dry stored

However, a more in-depth cleaning procedure (CIP) can be performed if the recommended cleaning above did not lower the initial TMP (transmembrane pressure) due to fouling. This deeper cleaning (CIP) involves soaking the membranes in chemicals for a longer time, 24 to 48 h. Here, the membranes are submerged in a chemical solution. The membranes can be soaked in hypochlorite acid, sulfuric acid and sodium hydroxide depending on the type of fouling on the membranes.

Initially, hand-held sprinklers were used, but they did not provide adequate pressure, resulting in ineffective cleaning and no significant reduction in TMP. Toward the end of the trial, new sprinklers capable of spraying at 2 bar pressure were introduced to better simulate full-scale operation. Cleaning was then carried out every 24 to 36 h, as recommended in the system design.

2.4 Sampling and Laboratory Analysis

Samples were collected from four specific sampling points, as outlined in Table 4, while the comprehensive analyses conducted are listed in Table 5. These analyses were carried out on random samples.

The water quality was analyzed before and after filtration through the ceramic membranes (MF FEED and MF PERM), and both microbiological and chemical analyses were performed. Two sampling points were used for the feed to the MF to ensure that there was no difference between the quality in the MF tank (MF FEED) what the membranes were actually filtering, and the dirty BW tank (MF RAW), which collects the UF1 dirty backwash. Flow cytometry (FC) analysis was performed on the permeate and compared to the water quality from the drinking water treatment plant. Total solids analysis was performed on the dirty BW and the sludge in the membrane tank.

Table 4. Defined sampling points.

Sampling point	Description
MF RAW	Dirty BW tank
MF FEED	Incoming dirty BW water to the membrane tank
MF PERM	Permeate, clean filtered water
MF SLAM	Sludge from the membrane tank

Table 5. Overview of the analyses performed during the trial period.

Microbiological analysis	Unit	Chemical analysis	Unit
Coliform bacteria	mpn/100 ml	Turbidity	FNU
E.Coli	mpn/100 ml	Color	mg Pt/l
Culturable microorganism	22 °C 3 days cfu/ml	UV-abs	254 mn 5 cm cuvette
Slow-growing bacteria	22 °C 7 days cfu/ml	pH	a.u
		Conductivity	mS/m
		Alkalinity	mg HCO_3/l
		Total organic carbon (TOC)	mg/l
		Dissolved organic carbon (DOC)	mg/l
		Suspended material (susp)	mg/l
		Aluminum, Al	mg/l
		Iron, Fe	mg/l
		Manganese, Mn	mg/l
		Total solids (TS)	g/l

The analyses were carried out at the accredited Norrvatten laboratory located at Görvälnverket, except for the total solids analysis, which was performed by an external laboratory, Eurofins AB. Samples for laboratory analyses were taken once a week from April to June 2025, resulting in eight chemical and microbiological analyses. For the total solids analysis, a total of five samples were sent to the external laboratory during April and May. The results are presented in Chapter 4.

2.5 Data collection

Data was collected using an online monitoring tool that measured flow rate (in liters per hour, l/h, equivalent to flux) and transmembrane pressure (TMP, in mbar). Negative TMP values indicate pressure across the membranes, while positive values represent backwash, during which water flows in the reverse direction. The online measurement records multiple measurements per minute, and therefore, each reported data point for flux and TMP represents the average value over one minute. Additionally, the total volume of filtrated water (m^3) and the average volume filtrated per minute were calculated.

The transmembrane pressure (TMP) was temperature-corrected to a standard reference temperature of 25˚C using the following equation:

$$Normalized\ TMP\ (25\ ^oC) = \frac{TMP_{Initial}}{1{,}024^{(25-T)}} \qquad \text{(Eq. 1)}$$

Where, $TMP_{Initial}$ is the measured TMP at the actual temperature T (in ˚C)

3. Raw Water Quality

The quality of the dirty backwash water from the UF-pilot significantly impacted the MF-pilot performance from late November 2024 to February 2025. The primary issue was insufficient retention time for the flocculation in the UF-pilot, resulting in small flocs in the dirty backwash water. This caused fouling on the MF-membranes, leading to shorter filtration runs. The main adjustment in the UF-pilot was to alter the retention time. In February 2025, the retention time was increased from 110 seconds to 180 seconds by installing a DN32 pipe. This resulted in slightly better flocs. However, in March 2025, the pipe was changed to DN50 while maintaining 180 seconds retention time, which resulted in larger and more stable flocs, thereby more representative period to evaluate the MF-pilot performance. Figure 3 shows the difference in flocs from the dirty backwash water during different operational settings in the UF-pilot and sedimentation qualities.

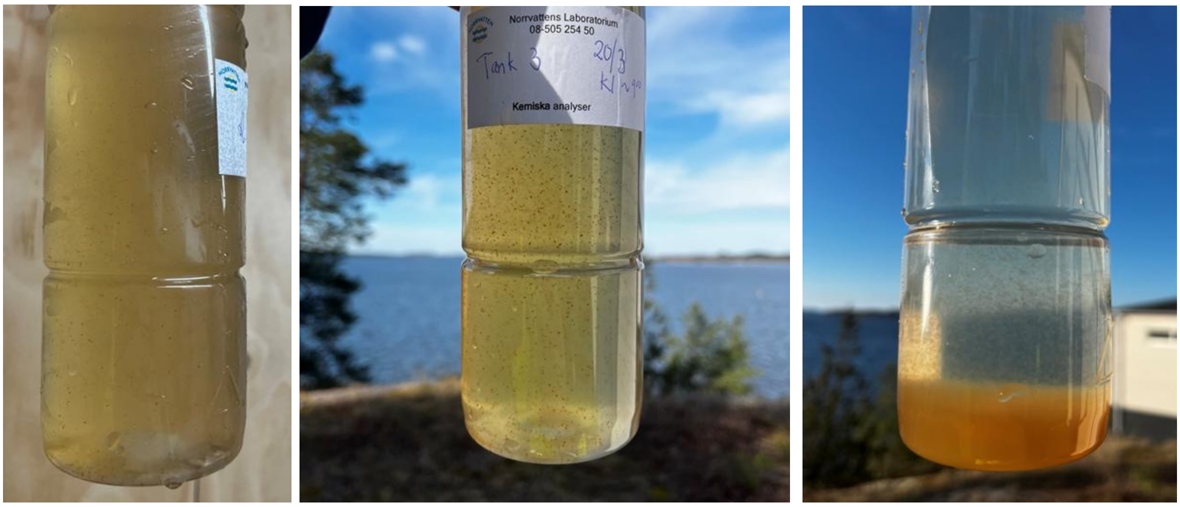

Figure 3. Left: show the dirty backwash water when the flocs are small and few (110 s flocculation); the flocs never sedimented, and the color remains yellow. Middle: shows clearer dirty backwash water when the retention time for the flocculation in the UF increased to 180 s. Right: shows the middle figure after 60 minutes of sedimentation.

4. Results and Discussion

This chapter presents and evaluates the results obtained during the membrane filtration trial. The dataset covers the period from March to June 2025, selected due to the integration of automation and online monitoring systems, as well as the most stable and representative operational conditions for assessing the membrane performance, as described in Chapter 3. Although this does not give a representation over the whole year, it was during a period of the spring lake turn, which is potentially challenging.

To evaluate membrane performance under higher loading, a flux of 150 lmh was briefly tested, along with adjustments to the backpulse interval. However, stable operation could not be achieved. Following this, the system was returned to 100 lmh, with backpulse intervals varied between 20 and 30 minutes to assess operational stability and cleaning effectiveness. A lower flux of 80 lmh was tested at the end of the trial period and the pilot was run overnight.

4.1 Pilot Run Performance Trends

This section summarizes the key performance trends observed during the pilot run. Figure 4 displays the entire trial period, with different cleaning procedures and flux rates indicated by color codes:

- Violet represents fluxs between 60-70 lmh. This range occurred overnight when the intended flux of 80 lmh dropped. This is due to generation of air bubbles in the permeate pump and sludge accumulation (requirement to de-sludge more frequently than 24 hours)
- Orange indicates a flux of 80 lmh
- Green indicates a flux of 100 lmh
- Navy blue indicates the highest flux tested, 150 lmh

Note that the pressure shown is suction pressure and is therefore negative. The filtrated volume refers to the total filtered volume through the membranes.

The colored lines described in Figure 4 represent a 30-point moving average, applied to highlight overall performance trends throughout the trial. The grey data points represent the raw, unfiltered data from which the moving average was derived. Figure 5 presents the same trial period as in Figure 4; however, in Figure 5, the TMP has been standardized to 25 ˚C, using Eq. 1.

The filtration trial was divided into 11 distinct periods, each corresponding to two cleaning cycles. From Period 7 to Period 11, the pilot was operated overnight, in contrast to earlier periods, where the system was shut down after daily operations; these overnight runs required a reduced flux setting (shown in violet) to ensure safe unsupervised operation. These overnight periods were further divided into sub-periods A and B, not due to cleaning events, but to indicate a change in flux settings within a single period.

The key operating parameters observed during the trial, as illustrated in Figure 4, are summarized for each period in Table 6. These include the targeted flux, filtration time, total filtrated volume, backpulse interval, approximate lowest and highest TMP during a filtration run, and the rate of TMP increase over time.

4.1.1 Trial with 100 lmh

The initial design flux was set to 100 lmh, period 1-3 and 5-6, the MF-pilot generally operated well. The backpulse interval was targeted at 30 minutes; however, during the trial, it was adjusted to between 20 and 25 minutes. The major issue encountered was the accumulation of air bubbles in the membrane permeate pump, which caused instability in maintaining a consistent flux at 100 lmh. When air entered the pump, the flux gradually decreased, necessitating degassing. Some challenges were also experienced with cleaning, primarily due to the quality of the dirty backwash water, which resulted in ineffective cleaning. However, soaking the membranes in a strong NaOH solution effectively restored TMP performance, likely due to the removal of biofouling.

4.1.2 Trial with 150 lmh

Following the trials at 100 lmh, it was considered that the MF-system might be capable of handling higher flux rates. A flux of 150 lmh was therefore tested, as this value aligned with the preliminary design specification provided by Cembrane for the second-stage membrane treatment at Norrvatten's planned new treatment plant. However, it was not possible to maintain a stable flux of 150 lmh for more than 10 to 15 minutes. This limitation was likely due to rapid fouling, although gas accumulation in the permeate pump may also have contributed. As shown in Figure 4, the 150 lmh trial (Period 4) was significantly shorter than the other test periods. Based on these results—and the inability to sustain a stable flow under these conditions, the flux was subsequently reduced back to 100 lmh.

4.1.3 Trial with 80 lmh

After testing both 100 lmh and 150 lmh, the membrane supplier recommended operating at a design flux of 80 lmh to maintain stable performance and enable longer filtration periods. In response, the flux was set to 80 lmh for the remainder of the trial. The backpulse interval was set to 20 minutes, which helped decrease the accumulation of air bubbles in the permeate pump and reduce the need for degassing. This setup was during period 7-11.

Despite automation being set to maintain 80 lmh, it was observed that the flux dropped to 60–70 lmh during overnight operation (During P5, see Table 3). This decline was attributed to the lack of monitoring, the inability to perform degassing, and sludge build-up in the membrane tank. A de-sludging or backwash was likely needed after approximately 16 hours of continuous operation.

At the end of this trial period, the membrane area was reduced to 0.5 m^2 with only three membranes plates in operation. This reduction was aimed at seeing if a more stable operation could be seen when the filtration area was built, and hence the feed flow was reduced. Overall, the operation of the plant was somewhat improved. However, the sampling of the permeate with the reduced membrane area revealed significantly higher values in metal analysis (Al, Mn, and Fe) and turbidity compared to when six membranes were in operation. The microbiological analysis indicated that the counts of culturable microorganisms and slow-growing bacteria were 5 to 10 times higher than those observed with six membranes in operation. This in all intense purposes due to poor sealing of the permeate ports to the three membrane plates that were removed. The aim of this period was not to investigate the quality, but to gain experience of longer operation of the system.

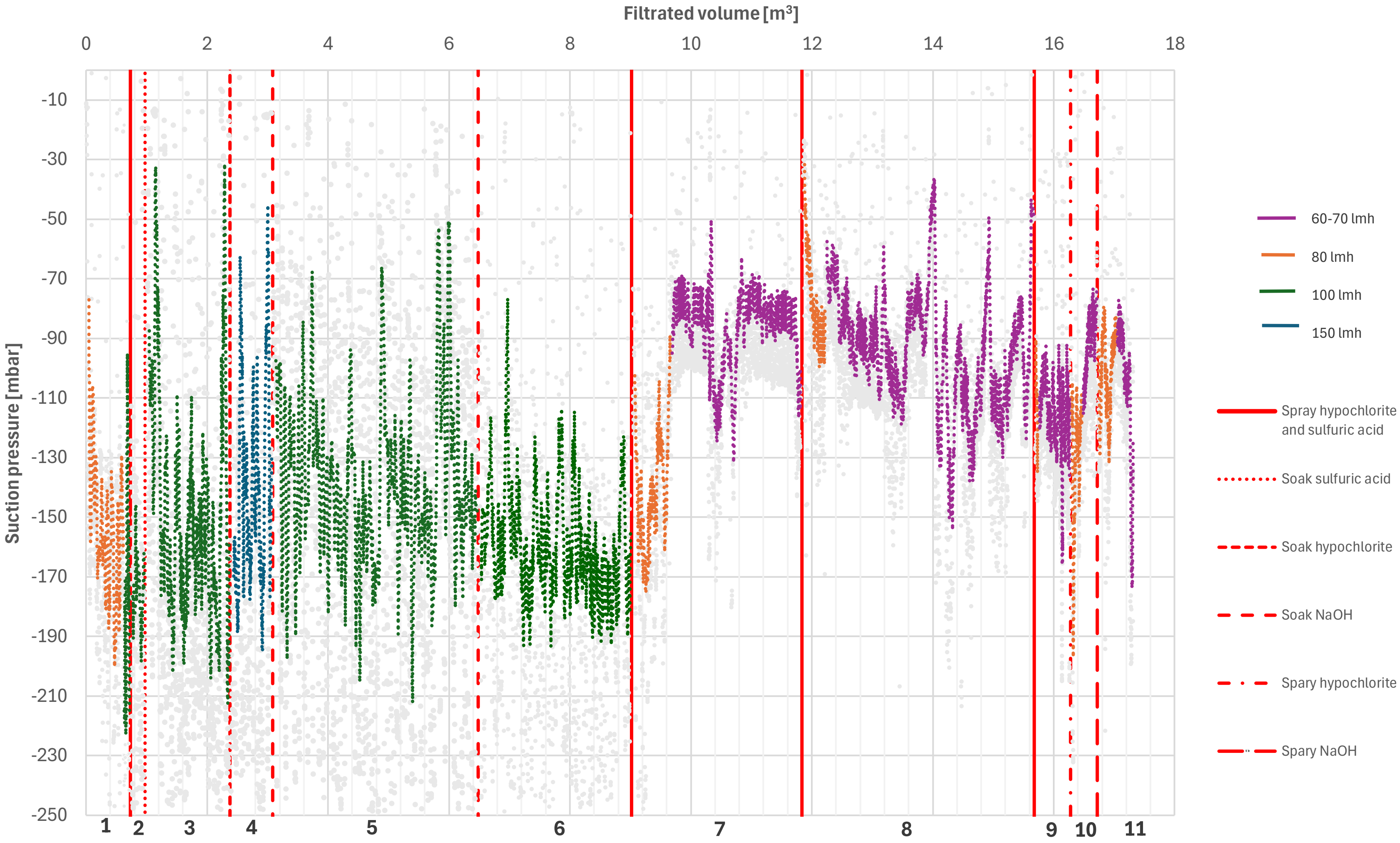

Figure 4. Trial period overview with numbered periods at the bottom of the diagram.

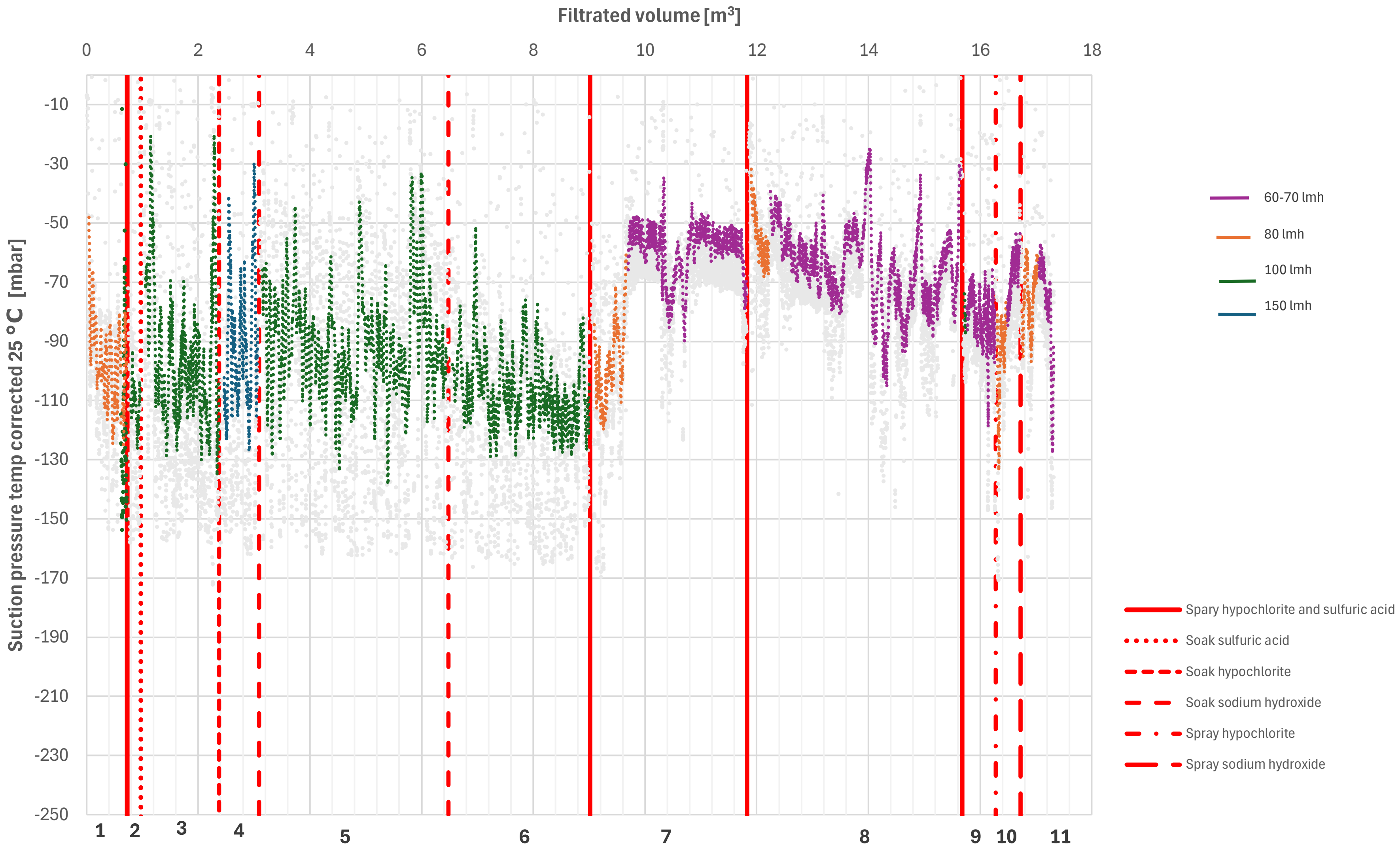

Figure 5. Overview of the trial period, where suction pressure is temperature corrected at 25 °C.

Table 6. Presenting flux, filtration run time, total filtrated volume, backpulse interval, the approximate minimum and maximum TMP for each run, both original and temperature corrected to 25 °C, and the build up over the membranes at 25 °C (TMP/min). A membrane cleaning was performed after each period.

Period	Flux [lmh]	Filtration run time	Filtrated volume (tot) [L]	Average temperature inlet BW [°C]	Backpulse interval [min]	Min and max TMP [mbar]	Min and max TMP at 25 °C [mbar]	Build up over the membranes at 25 °C [TMP/min]
1A	80	9 h 20 min	600	5	20-30	Min: -105 Max: -230	Min: -65 Max: -143	3.1
1B	100	1 h 30 min	100	5	20-30	Min: -115 Max: -230	Min: -70 Max: -143	2.9
2	100	1 h 40 min	200	6	20-30	Min: -115 Max: -215	Min: -73 Max: -137	2.6
3	100	17 h 10 min	1300	6	20-30	Min: -110 Max: -225	Min: -70 Max: -143	2.9
4	150	7 h 10 min	700	7	5–15	Min: -135 Max: -230	Min: -88 Max: -150	6.2
5	100	35 h 5 min	3400	7	20-30	Min: -95 Max: -220	Min: -62 Max: -144	3.3
6	100	26 h 30 min	2500	8	20-30	Min: -115 Max: -230	Min: -77 Max: -154	3.1
7A	80	7 h 20 min	600	9	20	Min: -105 Max: -175	Min: -72 Max: -120	1.9
7B	60-70	33 h 20 min	2200	9	20	Min: -80 Max: -105	Min: -55 Max: -72	0.85
8A	80	4 h 10 min	400	9	20	Min: -90 Max: -115	Min: -62 Max: -79	0.85
8B	60-70	56 h 15 min	3500	9	20	Min: -80 Max: -105	Min: -55 Max: -72	0.85
9A	80	1 h 10 min	50	11	20	Min: -105 Max: -130	Min: -75 Max: -93	1.1
9B	60-70	17 h 30 min	500	11	20	Min: -110 Max: -130	Min: -78 Max: -93	0.9
10A	80	4 h 40 min	200	12	20	Min: -115 Max: -150	Min: -84 Max: -110	1.3
10B	60-70	8 h 50 min	200	12	20	Min: -95 Max: -115	Min: -69 Max: -84	0.75
11A	80	7 h 50 min	300	12	20	Min: -95 Max: -120	Min: -69 Max: -88	0.95
11B	60-70	9 h 45 min	600	12	20	Min: -85 Max: -115	Min: -62 Max: -84	1.1

4.2 Chemical Cleaning Performance

During initial startup trials (late November to early February), the UF pilot experienced suboptimal performance during very low feedwater temperatures <5 °C. During this period, the UF1 backwash water appeared noticeably different compared to earlier UF testing, and the system showed elevated residual aluminum levels.

To address this, the flocculation time in UF1 was gradually extended, from 60 seconds to 120 seconds, and finally to 180 seconds. As a result, the UF pilot performance improved significantly, and the UF1 backwash water showed a much clearer flocs separation. Consequently, the MF performance improved, likely due to the improved feedwater quality.

To assess whether organic and biological contaminants had contributed to membrane fouling during the period when the UF pilot experienced issues with coagulant dosing, the membranes were initially soaked in NaClO for 24 h. This was followed by several filtration runs and then a second 24-hour NaClO soak. However, this procedure did not result in any noticeable improvement in membrane suction pressure. As a next step, the membranes were soaked in NaOH at pH 13 for 24 h, which successfully reduced the initial TMP. This NaOH soaking process was repeated twice, and it proved to be good enough in restoring membrane performance. In this context, soaking refers to submerging the membranes in drinking water within the membrane tank, followed by the addition of the cleaning chemical to reach the desired concentration. A summary of the various cleaning procedures and corresponding dates is provided in Table 4.

Table 7. Summary of the performed cleaning of the membranes.

Cleaning	Date	Type of cleaning
C1	March 31, 2025	Spray with NaClO (1000 mg/L) leave 1 h, followed by H_2SO_4 (approx. pH 2) leave 1 h
C2	April 1, 2025	Membranes soaked in H_2SO_4 (approx. pH 2) for 24 h
C3	April 7, 2025	Membranes soaked in NaClO (1000 mg/L) for 24 h
C4	April 14, 2025	Membranes soaked in NaOH (approx. pH 13) for 24 h
C5	May 6, 2025	Membranes soaked in NaOH (approx. pH 13) for 24 h
C6	May 14, 2025	Spray with NaClO (1000 mg/L) leave 1 h, followed by H_2SO_4 (approx. pH 2) leave 1 h. New sprinklers
C7	May 16, 2025	Spray with NaClO (1000 mg/L) leave 1 h, followed by H_2SO_4 (approx. pH 2) leave 1 h. New sprinklers
C8	May 23, 2025	Spray with NaClO (1000 mg/L) leave 1 h, followed by H_2SO_4 (approx. pH 2) leave 1 h. New sprinklers
C9	June 10, 2025	Spray with NaClO (1000 mg/L) leave 1 h. New sprinklers
C10	June 11, 2025	Spray with NaOH (approx. pH 13) leave 2 h. New sprinklers
C11	June 12, 2025	End cleaning before shutdown. Spray the membranes with NaClO (1000 mg/L) and H_2SO_4 (approx. pH 2). New sprinklers

Soaking the membranes in different chemicals had a positive effect on the start TMP. Soaking was carried out with H_2SO_4, NaClO and NaOH.

Later in the trial, the manual sprinklers were replaced with battery-powered sprinklers capable of delivering 2 bar pressure, significantly improving the effectiveness of chemical spraying with hypochlorite and sulfuric acid. This upgrade resulted in lower start TMPs after cleaning, compared to those achieved with manual sprinkling.

Ideally, the start TMP after each cleaning cycle should be lower than or equal to the previous cycle, indicating effective membrane recovery. However, it was observed that with manual sprinklers, the start TMP did not consistently decrease, likely due to insufficient spray pressure leading to ineffective cleaning.

4.3 Clean Water Flux Test

During the trial period, clean water flux test was periodically performed to benchmark the membrane permeability. To conduct this test, the membrane tank was filled with drinking water, and filtration was initiated at a low flux, which was then incrementally increased. For each flux setpoint, the TMP was recorded. The clean flux test for this trial was performed directly before and after a cleaning session. The result of the test is presented in Figure 6, the TMP has been standardized to 25 ˚C, using Eq. 1.

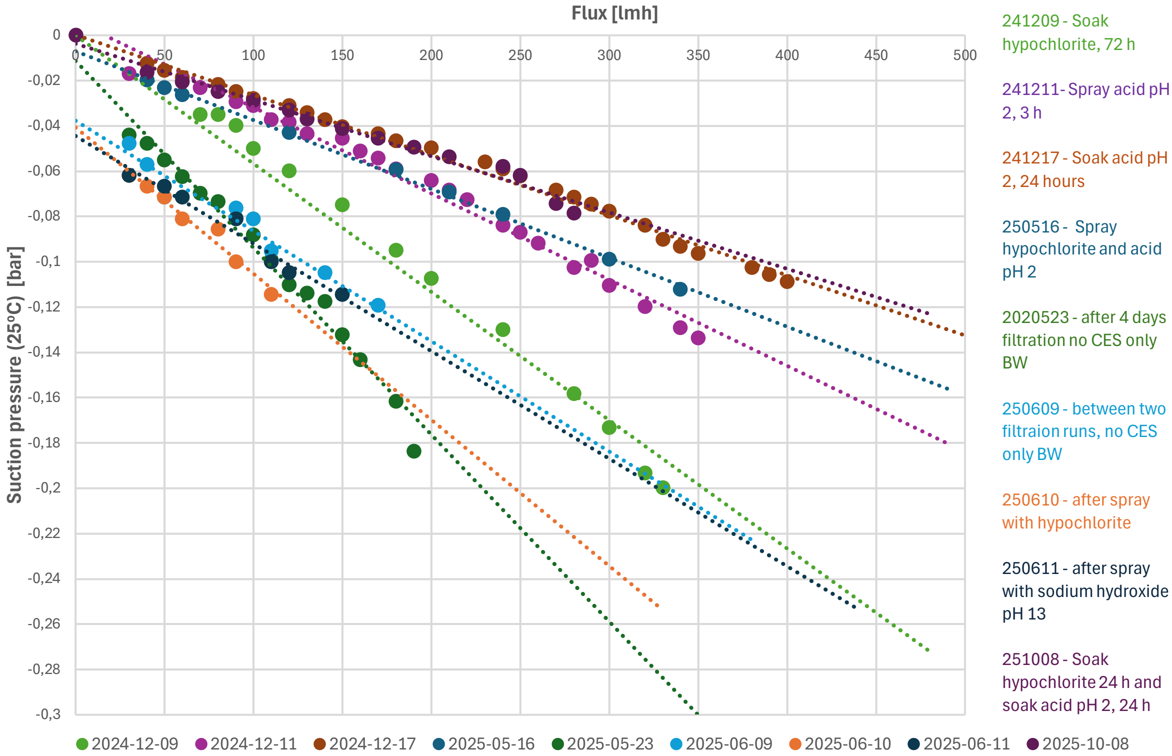

Figure 6. Clean water flux test at 25 °C.

As can be seen in the diagram in Figure 6, the clean water flux test results from before the trial period are displayed. These initial results were obtained during the start-up period and serve as a comparison with the most recent clean flux test, which was performed in the middle of October 2025. Table 8 below summarizes the clean flux permeability at different times during the testing period.

A key figure that has been calculated is the permeability (lmh/bar). New ceramic membranes from Cembrane should have a permeability of at least 3000 lmh/bar. As indicated in Table 8, permeability was higher during the start-up period compared to the end of the trial period, where there was a significant decrease in permeability. The cleaning from October was performed to examine the possibility of restoring the permeability of the membranes. Note that the pilot was not running continually when this cleaning was performed. The test performed in June 2025 was when the pilot was running daily.

Initially, the membranes had what is considered a poor clean flux permeability < 3000 lmh/bar (temperature corrected to 25°C).

During start-up in December 2024, the membranes appeared to be fouling quickly, this coincided with poor performance of the UF, which was struggling with low permeabilities and high aluminum levels in the filtrate. The reason for this was subsequently identified to be too low flocculation time at low feed water temperatures. The MF membrane also subsequently struggled with performance and most likely aluminum fouling. However, three cleaning cycles during December, before shutdown for Christmas recovered the clean water flux to > 3000 lmh/bar.
No clean water flux tests were completed between December and May 15 where a clean water flux test of 3000 lmh/bar was recorded, this shows that the cleaning cycles between December and May had been effective in recovering the membrane.

The clean flux test on June 10 showed there had been a significant loss of permeability. A subsequent clean on June 11 improved the recovery somewhat, but still 30% lower than the target of 3000 lmh/bar. The pilot was then stopped over the summer period due to resource issues. A more comprehensive and deep cleaning procedure with soaking the membranes in hypochlorite for 24 h followed by soaking the membranes in sulfuric acid pH 2 for 24 h showed an improvement in permeability, 3750 lmh/bar, see Table 8.

Table 8. Date description of the clean water flux test.

Date	Description	Clean flux (lmh/bar)
December 9, 2024	Startup period, after soaking the membranes in hypochlorite for 72 h	1767
December 11, 2024	Startup period, after spraying sulfuric acid pH 2 and leave for 3 h	2500
December 17, 2024	Startup period, after soaking in sulfuric acid pH 2 and leaving for 24 hours	3744
May 16, 2025	Spray with hypochlorite and leave for 1 h, and spray with sulfuric acid pH 2 and leave for 1 h	3000
June 10, 2025	After spaying with hypochlorite and leaving for 1 h (period)	1500
June 11, 2025	After spraying with NaOH pH 13 and leaving for 1,5 h	2000
October 8, 2025	Soaking in hypochlorite for 24 h followed by soaking in sulfuric acid pH for 24 h	3750

4.4 Critical Flux Test

Critical flux testing is conducted to determine the point at which membrane fouling deviates from a linear relationship between flux rate and suction pressure.

In this test the water to be filtered is pumped into the MF tank and then the permeate pump flow setpoint is set to a low start flux (e.g., 30-40 lmh). The flow and pressure are allowed to have a short stabilization period (approx. 1 min) and the pressure and flow are logged. The process is repeated with the permeate flow being increased in steps, until the pressure becomes too high or the pump cannot increase the flow.

This test should show a linear relationship between flow and pressure, but after a certain period, an inflection point will appear. This inflection point indicates the critical flux area/point where the build up of fouling is no longer linear. Operating at or just below the critical flux point should lead to a stable and robust operation. Operating above the critical flux is possible but will most likely require a more demanding design, having to do excessive backpulsing, backwashing and chemical cleaning.

During the start-up period, two tests were conducted in December 2024 and one more in May 2025 and October 2025. The test begins at approximately 10% of the target flux (here, 150 lmh), and the flux is successively increased by about 10%, while recording the corresponding TMP, until 150 lmh is reached. All critical flux test was performed after a cleaning cycle, see Figure 7, where cleaning was done before the test.

Figure 7 presents the results of the critical flux test. The TMP has been standardized to 25 ˚C, using Eq. 1. The critical flux lies in the marked grey area, within the interval of 80 to 120 lmh. This range is considered plausible, as sustaining operation at 150 lmh proved to be challenging. However, with proper automation, it is possible to reach the upper end of the interval, up to 120 lmh.

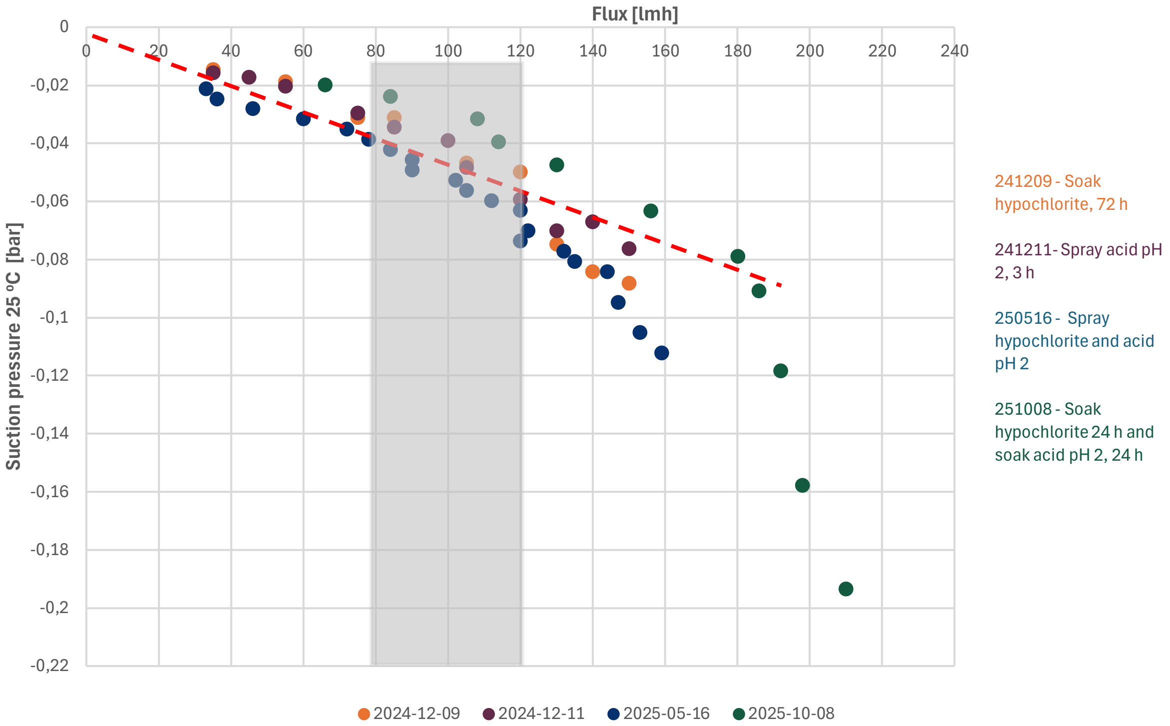

Figure 7. Presents the results of the critical flux test, temperature corrected at 25 ˚C.

4.5 Chemical and Microbiological Assessment

The chemical analysis, presented in Table 9, includes average values for samples taken from the dirty backwash tank (MF RAW), the dirty backwash water inlet to the membrane tank (MF FEED), and the permeate (MF PERM). Eight samples were collected once a week from April to June 2025.

In Figure 8, three water samples are shown to provide a visual representation of the analyzed water.

Figure 8. Left: Water from dirty BW tank. Right: Water from the inlet to the MF-pilot (MF Feed) and permeate water, sample after filtration (MF PERM).

Table 9. The average of eight sampling occasions for the chemical parameters before and after filtration.

Analysis	Unit	Average MF RAW	Average MF FEED	Average MF PERM	Removal [%]
Turbidity	FNU	43	55	0.41	99
UV-abs	254 mn 5cm cuvette	0.87	0.96	0.75	22
pH	a.u	7.1	7.1	7.1	-
Color	mg Pt/l		13	9.1	32
Conductivity	mS/m	25	25	26	-
Alkalinity	mg HCO_3/l		69	56	20
TOC	mg/l		62	7.7	88
DOC	mg/l		9,1	7.5	18
Suspended solids	mg/l		240	<2	>99
Al	mg/l		39	0.28	99
Al	acid soluble mg/l		0.049	0.016	67
Fe	mg/l		0.63	0.005	99
Fe	acid soluble mg/l		0.002	0.002	10
Mn	mg/l		0.12	0.002	98
Mn	acid soluble mg/l		0.001	0.001	41

The analysis indicates excellent removal of suspended solids and turbidity, as well as good removal of total metals. Interestingly, there is a slight reduction in dissolved aluminum, which could suggest aluminum adsorption or fouling on the membrane, an issue that the chemical enhanced soaking (CES) needs to address. Additionally, there is a small reduction in NOM (color, UV254), which could indicate potential organic fouling of the membrane—another concern that the CES needs to address.

There is a small reduction in alkalinity, which is most likely explained by the small pressure drop over the membrane, which can release a small amount of carbon dioxide and affect the carbonate balance.

Figures 9 and 10 illustrate the results from the microbiological analysis. Figure 9 displays the levels of coliform bacteria and E. coli, and Figure 10 shows the counts of culturable microorganisms (3 days and 7 days). The aim of performing microbiological testing was to evaluate and assess the quality and possible risks of recirculating the permeate back to the main treatment process.

Both the feed into the membrane tank and the permeate are visualized in these figures (9 and 10). Apart from the last sample taken on June 25, which had compromised port seal integrity, the MF membrane effectively removed E. coli and coliforms to non-detectible levels and achieved a very high percentage of removal for 3-day and 7-day plate counts. The June sample showed a high level of 3 and 7 day counts, a possible explanation for this is the backwash water used for backpulsing the membrane was taken from the treatment plant internal drinking water supply from a nearby main. This water contained a higher level of 3-day and 7-day plate counts compared to the MF permeate, especially when water temperatures increased in late spring, potentially allowing bacteria to transfer to the permeate and grow on the permeate side of the membrane. If a more readily available source of MF permeate were used for backwash, it is believed that these bacterial levels would be significantly lower. The average values for each analysis are provided in Table 10.

Note that Figures 9 and 10 include dotted lines as a representation to help visualize the variation in the count of bacteria from each sampling. As mentioned previously were eight microbiological samples taken, as the dots in the figures represent.

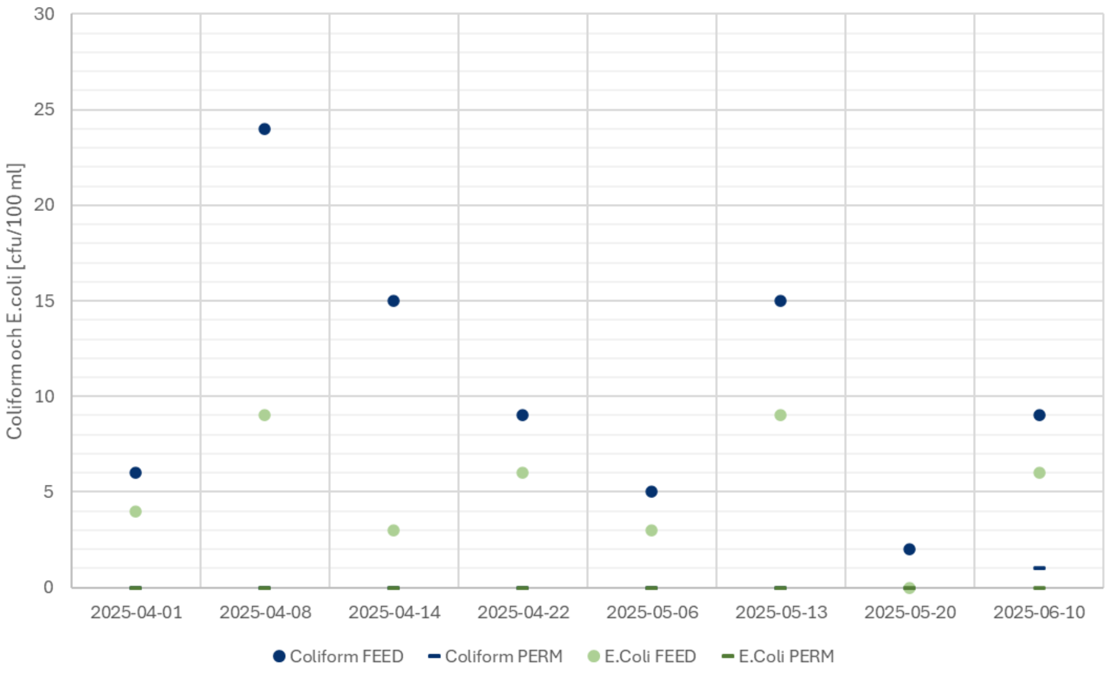

Figure 9. Illustrates the count of coliform bacteria and E. coli before (FEED) and after filtration (PERM), where results <1 cfu/ 100 ml is illustrated as 0 cfu/100 ml.

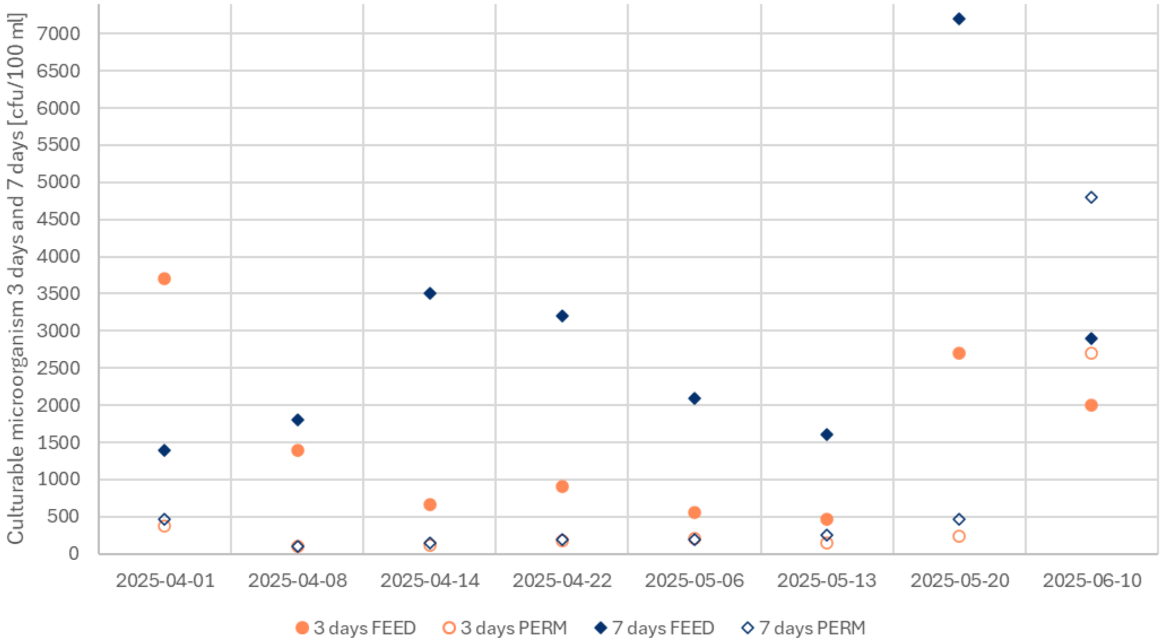

Figure 10. Illustrates the reduction of culturable microorganisms before (FEED) and after filtration (PERM).

Table 10. Average value of microbiological analysis.

Analysis	Unit	Average MF FEED	Average MF PERM	Removal [%]
Coliform	mpn/100 ml	11	<1	>91
E.Coli	mpn/100 ml	5.7	<1	>83
Culturable microorganisms, 3 days 22 °C	cpu/ml	1548	196	87
Slow-growing bacteria, 7 days 22 °C	cpu/ml	2963	263	91

Table 11 presents the results from flow cytometry (FC) analysis of the permeate. This analysis was conducted to characterize the cell population in the permeate and compare it with the drinking water produced at Görvälnverket as well as the permeate from the UF pilot. The MF drinking water refers to the drinking water used to perform backpulses and chemical solutions for the chemical enhanced soaking. The results show that the MF-permeate has significantly lower counts of total and intact cells, approximately the order of 2 log units lower than the outgoing drinking water from the treatment plant. This demonstrates that the MF system provides a high microbiological barrier, effectively treating source water with high levels of suspended solids, and elevated levels of bacteria and viruses compared to the raw water to the treatment plant.

Table 11. Results of flow cytometry (FC) tests from MF-Pilot.

Sampling date	Sample	TCC, Total cell count [10^3 cells/ml]	ICC, Intact cell count [10^3 cells/ml]	Intact cells [%]
2025-01-04	MF PERM	24	17	72
2025-04-08	MF PERM	13	11	89
2025-04-14	MF PERM	9.2	5.5	60
2025-04-14	MF Backwash water drinking water	1236	924	75
2025-04-14	Outgoing drinking water Görvälnverket	1239	1103	89

4.6 Sludge Assessment

The sludge analysis was performed at two sampling points using random sampling:

- Dirty backwash tank (DIRTY BW)
- Sludge from the bottom of the membrane tank (MF SLAM)

The sludge from the bottom of the membrane tank was collected via the drain valve, see Figure 11 for a visual representation of the concentrated sludge at the bottom of the membrane.

Figure 11. Sludge in the bottom of the membrane tank.

The results are shown in Table 12, presenting the measured total solids in the dirty BW tank (UF1 dirty backwash water) and the sludge collected from the bottom of the MF membrane tank. The initial two sludge samples from the membrane tank (MF SLAM) were collected at the start of the pilot tank drainage. A 500 ml bottle was filled with the first flush, which contained sludge from the bottom of the tank mixed with water above the sludge, resulting in a lower number of total solids, see Table 11. The last three samples were collected after the tank was drained and contained only the sedimented sludge (lower water level). Consequently, thicker sludge was collected, which is considered to be more representative of the sludge samples.

The thickest sludge in the membrane tank had a total solids value of 7.1 g/l and 8.0 g/l (approximately 0.7-0.8% TS), while the total solids in dirty BW tank were overall in the same range of 0.38 to 0.46 g/l for each sample.

Since the existing sludge treatment process at Görvälnverket will continue to manage the sludge from the new drinking water treatment plant, it is important to get an idea of the amount of sludge produced by the MF system. An estimation of the likely MF recovery was calculated from measuring the volume of sludge drained from the tank and comparing it to the permeate volume filtered through the membrane. This allows for the calculation of an estimated "Recovery over the MF", presented Table 12. Consequently, a sludge yield has been calculated.

Simplistically, if the MF recovery is > 90 %, the designed volume of sludge produced by a potential MF2 solution (using MF-membranes as a second treatment step for UF backwash water) in the NFVP project will be less than the volume that will exceed loading rates in the existing sludge treatment process. This is likely to cause exceedances in quality targets to the recipient water. As Table 12 shows, the recovery is above 90%, which indicates existing sludge treatment can handle the waste stream from the new process from NFVP project.

Table 12. Results of total solids for the dirty BW tank and MF Sludge.

Sampling date	Total Solids, DIRTY BW [g/l]	Total solids, MF SLAM [g/l]	Operate time before sampling [h]	Estimated water volume [l]	Estimated sludge volume [l]	Recovery [%]
2025-04-01	0.41	1.9	2.5	250	11	96
2025-04-08	0.39	2.3	4.5	675	30	96
2025-05-12	0.46	8.0	5.5	550	26	95
2025-05-20	0.42	7.1	20.5	1230	65	95
2025-05-21	0.44	7.9	17.5	875	56	94

5. Summary and Conclusion

The current proposed UF2 system in the NFVP project when operating at design capacity produces a volume of backwash waste to the existing sludge treatment system that increases the loading (including future extension) to a level in excess of the current loading rate which can lead to discharge limits for suspended solids potentially being exceeded.

Moving from a pressurized ultrafilter membrane to a submerged filter system allows for a system which can handle high levels of feed suspended solids, produce similar permeate water quality to UF2 process, and increase the recovery, thus reducing the amount of waste/sludge that will be pumped to the existing sludge treatment system. Hence, meet acceptable loading rates in the existing sludge system and subsequently meet discharge quality requirements.

Flatsheet ceramic microfilter membranes are a relatively new membrane technology that can offer advantages over polymeric membranes. The membranes often can operate at higher flux rates to the polymeric membranes and are not limited in cleaning chemicals and have a much longer lifetime. However, they do come at a higher cost than polymeric membranes. Additionally, very little reference information exists for the operation of flatsheet ceramic membranes in the recovery of chemically coagulated wash water, thus necessitating this trial.

Ideally, a full-scale automated MF-pilot plant would have been preferred. However, the feed requirement to the smallest of these systems was well in excess of the amount of backwash waste being produced by the UF1 pilot already in operation at Norrvatten. Split Nordic Water, together with Cembrane proposed a mini-pilot that was not a fully automated pilot system; it was a simpler bench-scale system to help identify if ceramic membranes are a good potential option to second-stage ultrafiltration in the NFVP project. The mini-pilot's aim was to identify the potential design flux and operating conditions for an MF-ceramic membrane system for NFVP.

Operation of the mini-pilot proved challenging; it required a significant resource input, had problems with the build-up of gas in the feed pump, and had limited data logging technology. However, with the assistance of Split Water Nordic an amount of automation was added to the system, which reduced the resource input and provided data logging. From March 2025 useful data could be gathered from the mini-pilot to allow conclusions to be made, however, with some caveats.

Tests showed that a design flux in the region of 80-120 lmh should be possible with a ceramic MF solution at a recovery rate of 90-95% which meets the threshold of the existing (upgraded) sludge waste system. The membranes appeared to show good cleanability, however, at the end of the trial the clean water permeability had dropped somewhat even with the normal cleaning regime but was recovered back to the original level with a deep clean with longer soak times. Any future design needs to allow redundancy for this longer cleaning over 1-2 days. Further testing with a fully automated pilot needs to be completed to confirm how often these deep cleaning or CIPs need to be done.

The quality of the permeate from the ceramic membrane cleaning in question and can be considered similar to a UF polymeric membrane. Although the ceramic has a slightly larger pore size than UF-membrane, cake layer filtration due to the high solids loading creates an extra filtration effect on the MF-membrane. Test showed high levels of rejection of cells using flow cytometry, a significant improvement when compared to the existing drinking water produced at the treatment plant, even though it was treating a significantly higher solids and microbial load.

The mini-pilot was not properly able to demonstrate how the membrane would perform over a continuous 24/7 period due to the limitations mentioned earlier. Thus, making it difficult to propose a process solution without further testing.

The ceramic MF-membrane has shown good potential for handling high suspended solids generated by the UF pilot backwash waste, at a potential flux rate of 2-3 times of a proposed pressurized secondary UF membrane system. This needs to be balanced though with higher price of ceramic membranes, which are of the order 10 times more than polymeric membranes (20–30 €/m^2 polymeric compared to 250-300 €/m^2 ceramic membranes). However, the lifetime of ceramic membranes is considerably longer, and they are potentially much more suited to handling high solids and less prone to irreversible fouling compared to pressurized polymeric membranes.

Using a conservative design flux of 80-90 lmh, preliminary designs have shown that a ceramic MF- membrane treating the first stage UF backwash waste can be installed in the allocated area for the earlier proposed UF2 system in project NFVP.

The risk of a ceramic membrane plant with a design flux of 80 lmh and one redundant MF filter not achieving the project's design requirements is considered minimal. If a ceramic MF solution is chosen as the UF 1 washwater recovery process, a fully automated ceramic MF pilot with a full-sized module should be included in the planned future pilot to validate and optimise the full-scale design before installation.