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“Interstage crystallization to enhance two stage reverse osmosis (RO) recovery”

Milestone 6: Final Report

Peter Sanciola

Eddy Ostarcevic

Paul Atherton

Greg Leslie

Tony Fane

Yoram Cohen

Marrack Payne

Stephen Gray

Institute for Sustainability and Innovation

Victoria University

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Glossary

ASP : Accelerate seeded precipitation, a process in which precipitation from supersaturated solutions is accelerated by adding a seed material that acts as a nucleating site for precipitation.

Brackish water: Water that contains between 0.5 g/L and 30 g/L of salt

CaHPO₄ (Calcium hydrogen phosphate): The chemical formula of the solid formed when large quantities of phosphate ion are added to RO concentrate in this trial

Calcium selective electrode: A sensor that converts the concentration of calcium ions dissolved in a solution into an electrical potential which can be measured by a voltmeter or pH meter. The sensing part of the electrode is of membrane which only allows calcium to cross the membrane

Ceramic Ultrafiltration (CUF): Ceramic membranes which typically have pore sizes in the range of 0.01 - 0.10 μm (i.e. 0.01 to 0.1 thousandths of a millimeter), and are resistant to harsh chemical cleaning conditions (e.g. very concentrated acid for long periods)

Dissolved air flotation (DAF): A process of separation of solids from solution, that relies on the formation of small bubbles within particle aggregates, causing the particles to rise to the surface and allowing removal of these particles using a skimming device

Equimolar: Having an equal number of moles

LSI: Langelier Saturation Index. This is a measure of the tendency for calcium carbonate to precipitate. = pH of water – pH at which the water is saturated with respect to calcium carbonate.

Municipal Wastewater: Effluent from wastewater treatment plants that deal with sewage from residential and industrial sources

Na₂HPO₄ (Disodium hydrogen phosphate): A solution of this salt was used in this trial to increase the concentration of phosphate in the RO concentrate in order to precipitate it as calcium hydrogen phosphate (CaHPO₄)

pH: A measure of the acidity of solutions, $\text{pH} = -\log_{10}[\text{H}_3\text{O}^+]$

Reverse Osmosis (RO): Filtration process using membranes which typically have pore sizes less than 0.001 microns, and allow water to move through the membrane under the influence of an applied pressure, but do not allow salt ions and organic materials to pass through

Sedimentation: A process for separation of solids from liquid, that relies on the solid's ability to form a sediment under the action of gravity

Ultrafiltration (UF): Filtration process using membranes which typically have pore sizes in the range of 0.01 - 0.10 μm (i.e. 0.01 to 0.1 thousandths of a millimetre)

Executive Summary

The prospect of decreasing the requirement for evaporation ponds, and the generation of more clean water from a limited municipal wastewater resource, at a time when access to water for agricultural enterprises in rural locations is likely to decrease, is a very attractive prospect. Calcium scale formation in reverse osmosis equipment limits the quantity of clean water that can be extracted from a wastewater source, and results in a significant quantity of concentrated waste brine for disposal. This project investigates the feasibility of removing calcium scale precursor ions from RO concentrate, allowing further RO treatment of the remaining concentrate to extract more water, thereby increasing the overall water recovery and decreasing the quantity of waste brine.

The test water in this investigation was wastewater from a wastewater treatment plant that treats the municipal waste from Donald – a small town with population of 1400 in the Grampians Wimmera Mallee region of Victoria. This wastewater is characterised by its relatively high salinity (10,000 mg/L) and calcium content (220 mg/L), which are believed to be influenced by high evaporation due to low throughputs, and infiltration of brackish groundwater. Due to the low flows, this wastewater is also characterised by high turbidity (300-500 NTU) due to algal growth. Laboratory testing in the early stages of this project was largely focused on the chemistry of accelerated seeded precipitation of calcium ions from the wastewater. The process was tested on a larger scale in a pilot plant trial in Donald, Victoria, with a 25 kL/day pilot plant consisting of dissolved air flotation, ultrafiltration, reverse osmosis, and accelerated seeded precipitation. A ceramic filtration system was used to remove the seed and precipitated calcium and return the seed to the reaction tanks, and to supply a calcium depleted stream for further water recovery via RO.

Laboratory trials

Laboratory trials in the early stages of this project have shown that the removal of calcium ions from the test water by accelerated seeded precipitation (ASP) using calcium carbonate seed was not effective. This was attributed to the presence of substances such as magnesium ions and phosphate ions which interfere with calcium carbonate precipitation. The precipitation of calcium as calcium phosphate, however, was found to be less prone to interference. In-situ generation of seed by addition of phosphate ion at pH 10, followed by reuse of this seed, was found to be an effective means of removal of the calcium. The quantity of phosphate required in these laboratory experiments was one half to one third of the quantity (in moles) of calcium.

Pilot Plant Trial

The primary objectives of this research were to operate the ASP process at pilot scale in a continuous manner and integrate it into an RO process to achieve high water recoveries, determine the chemical and energy requirement of the process and perform preliminary cost estimates.

The pilot plant used in these studies consisted of dissolved air flotation, ultrafiltration, reverse osmosis, accelerated seeded precipitation and ceramic ultrafiltration. The large number of process steps in the final plant design made uninterrupted operation of the full plant for more than a few hours very difficult, resulting in a much longer than planned commissioning period. Approximately 5 of the six months of this trial was employed for the commissioning process. Despite this major difficulty, however, the plant was operated for long enough for the major impediment to its operation and to the prolonged achievement of high water recovery to manifest itself – turbidity

forming after the ceramic ultrafiltration (CUF) system that prevented further RO processing to extract more clean water.

Running of the pilot plant was characterised by periods of extended low turbidity (<1 NTU) in the CUF filtrate followed by unexpected increases in turbidity above the 1 NTU requirement for the RO process. The generation of a turbid CUF filtrate was tentatively attributed to the heating in the ceramic filter recirculation loop due to pump energy input, giving rise to the generation of a supersaturated calcium solution due to the decreased solubility of sparingly soluble calcium salts at the higher-temperature, and the precipitation of calcium from this supersaturated solution after it passes through the ceramic filter. The increases in CUF turbidity during this trial necessitated the disconnection of the CUF filtrate from the RO, thereby reducing run time and preventing investigations which require longer time periods, such as seed dose and chemical optimisation, and RO scale formation assessments.

Preliminary cost estimates for the ASP process indicate that the ASP-RO process is considerably more expensive than conventional RO, but that there are considerable cost savings associated with lower evaporative pond required to manage the waste brine. The chemical running costs associated with the ASP stage suggested from the laboratory experiments were \$1.8 per kL of product water. The un-optimised pilot plant ASP stage chemical usage costs were \$20 per kL of product water. The energy running costs of the plant were estimated to be \$3.8 /kL of product water, which is 13 times the energy cost of conventional RO run at 75% water recovery. At an equal product water output, the cost of evaporation pond construction at 90% water recovery were estimated to be one third those of conventional RO run at 75% water recovery.

Recommendations for Further Research

The removal of the seed and precipitated calcium is central to the success of the process, and further research is required to determine the best method of removal of these solids. The chemistry of calcium phosphate precipitation appeared to vary throughout these trials because of changes in temperature, and thus the design of the solid-liquid separation process in the ASP was unable to achieve the required low turbidity values under all conditions. The kinetics of calcium phosphate precipitation as a function of temperature and the mode of generation of this turbidity needs to be investigated. It is only after an effective and reliable mode of separation of the seed and precipitated calcium salts from the treated RO concentrate is found that the seed material can be allowed to build up to useful concentrations (~20 g/L according to laboratory tests) and optimisation of process inputs such as chemicals and energy can be made. Other modes of solids separation, such as sedimentation, should also be considered for this application. Sedimentation is a more conventional approach but was originally considered less flexible for these small scale pilot tests.

The detection of the incidence of scale formation relies on comparisons of flux and pressure that are best performed at constant feedwater composition. The recirculation of the calcium-depleted ASP-treated RO concentrate back to the same RO that had generated the concentrate in this trial resulted in long periods during which the feed water to the RO was changing in composition, making early detection of scale formation very difficult. Further research in high recovery RO would be facilitated by the use of two smaller RO plants in series rather than one larger RO unit operated in recycle mode.

It would be desirable that future research projects in this area should focus exclusively on the ASP, solids removal and downstream RO processing, without having to in any way be concerned with pre-treatment and the generation of the RO concentrate that is to be feedwater for the ASP process. A willing partner that already performs the pre-treatment and generates a concentrate should be sought and the pilot ASP research facilities should be set up at the source of the RO concentrate that is to be feedwater for the ASP process. The cost of chemicals and energy for this research can be minimised by constructing and operating a smaller ASP plant.

Although optimisation of the chemical and energy inputs was not performed during this trial, it is obvious that the process is likely to be an expensive one for the highly saline (~10 g/L salt) and calcium rich (~0.2 g/L) test water used in this trial. Future research should consider application to less saline and calcium rich wastewaters that would require less chemical and energy inputs.

Potential benefits of high recovery desalination to local water Industry

Addition of another process stage to the conventional RO desalination process to achieve higher water recoveries is expected to incur an added cost associated with the additional chemicals and energy inputs. The major benefit of this added cost, however, is that the quantity of waste brine produced is less. At equal clean water production rates, the evaporation pond requirement for a plant operated at 90% water recovery is one third that of one operated at 75% water recovery. If 95% water recovery could be achieved, the evaporation pond requirement would be reduced to one sixth that at 75% water recovery. This would, however, require considerably more energy than operation at 90% water recovery, and much more than operating at 75% water recovery.

Economic justification for aiming to achieve higher water recoveries can only be made if the benefit of the lower evaporation pond requirement outweighs the higher chemical and energy running costs. This is most likely to be the case in locations where there is a desperate need for fresh water and where land costs and regulatory requirements make the construction of large evaporation ponds prohibitively expensive.

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1 Introduction

1.1 Limits to Reverse Osmosis (RO) process water recovery

Generally, conventional RO treatment of brackish water can only recover between 55% and 85% of the wastewater (1) for reuse, leaving a large volume of waste saline brine requiring costly disposal and representing a waste of water resource. Water recoveries greater than these usually result in the formation of calcium scale on the membrane surface, resulting in an increase in energy and cleaning requirement, and a decrease in the lifetime of the RO membranes. This project investigates the feasibility of removing the calcium scale precursor ions from the RO concentrate via precipitation, decreasing the likelihood of calcium scale formation and allowing further RO treatment of the remaining concentrate to extract more water, thereby increasing the overall water recovery.

1.2 Seeded precipitation theory and practice

The extraction of low salt water in the RO process leaves behind a waste brine that contains the vast majority of the salts originally present in the feedwater. The concentration of these salts depends on the water recovery of the process, and often results in a metastable calcium solution – a salt solution which is supersaturated with respect to sparingly soluble calcium compounds such as calcium carbonate, calcium phosphate or calcium fluoride. In such a situation, the precipitation of these calcium compounds can be triggered by the addition of solid particles that act as nucleating sites for precipitation. These solid particles are often chemically identical to the calcium compounds to be precipitated. The exact requirements of the seed material to be used are not clear from the literature on the subject, but it is often found that similarity with the compound to be precipitated is beneficial.

1.3 Novelty of the current research

Previous work in this area has mainly dealt with high recovery treatment of surface waters using calcium carbonate (2) or calcium sulphate seed (3) to precipitate calcium carbonate or calcium sulfate, and the separation of seed from the treated water via sedimentation. The current research applies the concept of calcium removal via precipitation to achieve higher RO water recoveries to the treatment of saline municipal wastewater. Due to the presence of wastewater constituents (HPO_4^{2-} , Mg^{2+}) that interfere with the accelerated precipitation of calcium carbonate, calcium phosphate seed was used.

Prior research in this area has also used sedimentation clarifiers and filtration to remove the seed from the treated brine prior to further RO treatment. Clarifiers require a large footprint, are prone to problems associated with gas bubble formation and are limited to low solids loading. Sedimentation is not suited to high solids loadings as interaction between the particles can slow sedimentation. The current research used a ceramic filtration system that promised to be more reliable and have a smaller footprint than a clarifier. The use of a ceramic filter rather than sedimentation clarifiers also had the added advantage of allowing operation at higher solids loadings and thereby the potential to lower phosphate ion additions. Ceramic ultrafiltration was used for this application due to the ability of ceramic membranes to withstand the harsh cleaning required to remove calcium phosphate fouling. It is, however, much more energy intensive as mitigation of fouling due to cake formation is

accomplished with the use of very high flows, rather than backwashing as is the case for conventional ultrafiltration with organic membranes.

1.4 Laboratory experiments

Calcium carbonate, calcium sulphate and calcium phosphate seeds were trialled on Donald RO brine that had been produced at 70% water recovery. Calcium sulphate was found to be unsuitable due to its high solubility. Calcium carbonate was found to be unsuitable as very large quantities of calcium carbonate seed (~160 g/L) were required to reduce the calcium concentration to one tenth (~25 mg/L) of the original concentration. Calcium phosphate seed was found to result in low calcium concentration (~11 mg/L) at a dose of 10 g/L.

Calcium phosphate is considerably more expensive than calcium carbonate, so the feasibility of removal of phosphate – a known interfering species in calcium carbonate precipitation – was investigated by addition of aluminium ions and filtration prior to the seeded precipitation using calcium carbonate seed. This pre-treatment did not improve the seeded precipitation using calcium carbonate. This was attributed to the high concentration (370 mg/L) of magnesium ion - another powerful inhibitor of calcium carbonate precipitation – in the wastewater.

The feasibility of generation of the calcium phosphate seed in-situ via the addition of phosphate ion was also investigated. It was found that increasing the molar concentration of phosphate to that of the incoming calcium was an effective means of generating the seed material and decreasing the calcium concentration. Filtration of the generated seed and re-suspension in untreated RO concentrate was found to be an effective means of decreasing the calcium concentration. It was, however, found that repeated reuse of the in-situ generated seed decreased its effectiveness. This was attributed to the fouling of the seed surface by surface precipitation of other calcium and magnesium compounds (e.g calcium carbonate, magnesium carbonate). The required phosphate concentration for effective calcium removal using reused in-situ generated seed was found to be one half to one third of the calcium concentration (in moles/L).

1.5. Trial location and process feedwater

The trial was conducted in the wastewater treatment plant that treats the municipal waste from Donald – a small town with population of 1400 in the Grampians Wimmera Mallee region of Victoria. This wastewater was characterised by its relatively high salinity (10,000 mg/L) and calcium content (220 mg/L), which are believed to be influenced by infiltration of brackish groundwater. Due to the low flows, this wastewater was also characterised by high turbidity (300-500 NTU) due to algal growth. Treating this wastewater by conventional RO would be deemed impractical as the high calcium content would necessitate the operation at relatively low water recoveries (73% for an LSI of less than 1, 81% for an LSI of 1.5, at pH 6.5), generating large quantities of waste brine which would require large evaporation pond areas to manage.

1.6 Specific objectives

The primary objectives of this research were:

- a. To operate the ASP process at pilot scale in a continuous manner and integrate it into an RO process to achieve high water recoveries;
- b. Determine the chemical requirement of the process;
- c. Evaluate calcium selective electrode as a real-time monitoring method for Ca;
- d. Identify evidence of fouling on the RO membranes;
- e. Identify if fouling of the ceramic UF is significant over the life of the trials, and
- f. Perform preliminary cost estimates.

2 Method

2.1 Trial plant

The trial plant train consisted of pre-treatment using dissolved air flotation (DAF) and ultrafiltration, followed by desalination using reverse osmosis, accelerated seeded precipitation and ceramic ultrafiltration as shown in Figure 2.1.1.

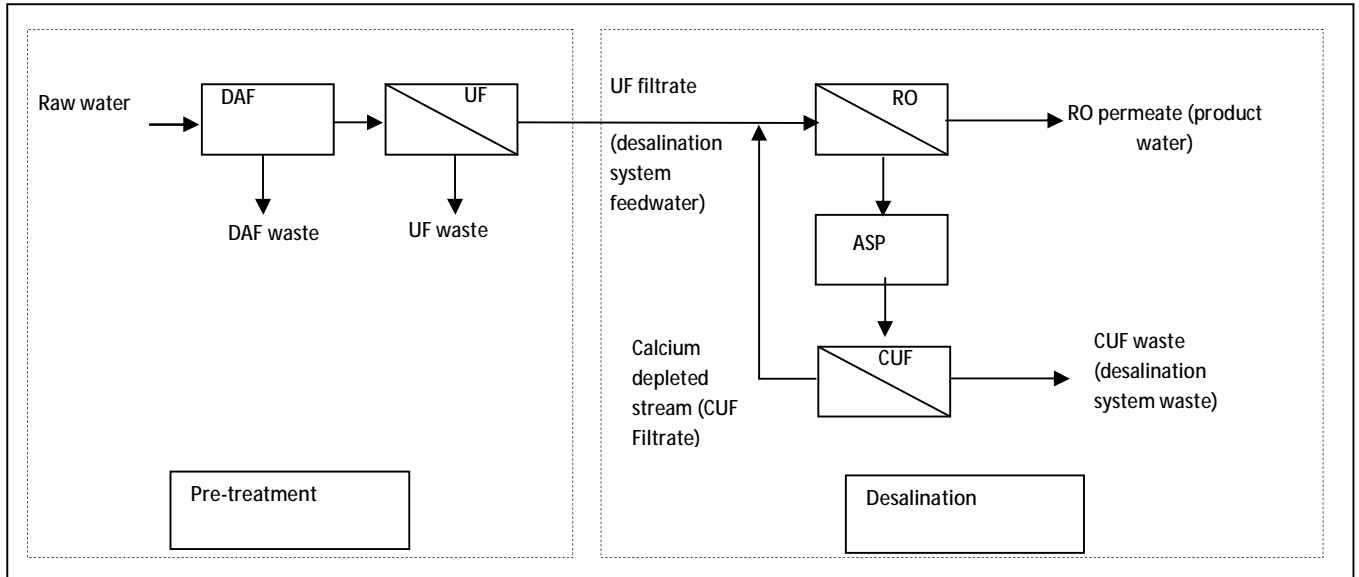


Figure 2.1.1: Diagrammatic representation of process stages and streams. DAF: Dissolved Air Flotation; UF: Ultrafiltration using Hydracap membrane modules; RO: Reverse osmosis using Filmtec SW membrane elements; ASP: Accelerated seeded precipitation using pH elevation and phosphate ion addition; CUF: Ceramic Ultrafiltration.

DAF plant component: Removal of algae from the municipal wastewater feedwater was performed using a DAF plant (1.5 m x 1.0 m x 1.0 m). Ferric chloride was used as coagulant at 50 to 100 ppm dose rate, depending on turbidity of the feedwater. This treatment was found to reduce the turbidity by approximately 50 to 60%.

UF Plant Component: The feedwater to the RO was further pre-treated by ultrafiltration using 4 HYDRAcap 60 (Hydranautics) membrane capillary (inside out) modules. This treatment was found to reduce the turbidity to below 1 NTU.

RO Plant Component: Reverse osmosis was performed using Dow,- Filmtec membrane elements (6 x 4 inch diameter seawater elements (SW4040), 6 x 2.5 inch diameter seawater elements (SW2540), in series configuration).

ASP-CUF Plant Component: 100 L flash mixer, 200 L flocculation tanks, 200 L maturation tank. Addition of caustic soda was performed in the lines to the flash mixer and flocculation tanks. Filtration of the precipitated calcium salts from the brine was performed using Jiangsu Jiuwu Hitech Co., Ceramfil® Inorganic Ceramic Membranes consisting of multi-channel tubular elements composed of alpha-Alumina/Zirconia with a 50 nm pore size (Model CMF09060-OD25), assembled in 3 pressure vessels, each pressure vessel containing 19 ceramic membranes. The total surface area of the ceramic filtration system was 11.4 m². The ceramic filter system was equipped with an internal

solids recycling loop designed to minimise caking by providing a high flow rate within the ceramic filter elements (typically 70,000 L/h for a filtrate flow of approx. 800 L/h). The filtration system was also equipped to return the ceramic filter concentrate to the ASP flocculation tanks, allowing higher seed dose in the flocculation tanks.

Analytical techniques: On site monitoring of calcium levels was performed using a calcium ion selective electrode. Validation of this method on the test waters prior to the trial was performed by comparison of the calcium selective electrode results with those obtained using atomic absorption spectroscopy.



Figure 2.1.2: Pilot Plant component, DAF (top left), UF (top right), RO (bottom left) and ASP-CUF (bottom right).

2.2 Trial operating conditions

The system operating conditions used in the pilot plant trials are shown in Table 2.2.1.

Table 2.2.1: Run conditions

	1	2	3	4	5	6		
Run Condition (RC) No.	UF filtrate flow (kL/d)	RO feed flow (kL/d)	RO permeate flow (kL/d)	RO concentrate flow (kL/d)	Calcium depleted stream flow (kL/d)	System waste flow (kL/d)	System water recovery (%) (RO permeate/UF filtrate) x 100	RO water recovery (%) (RO permeate/RO feed) x 100
RC1	25	50	22.5	27.5	25	2.5	90	45.0
RC2	37.5	50	33.7	16.2	12.5	3.7	90	67.5
RC3	37.5	50	35.2	14.7	12.5	2.2	94	70.5
RC4	41	50	36.9	13.1	9	4.1	90	73.8
RC5	19.3	36.3	17.46	18.9	16.9	1.9	90	48.0

The purpose of run conditions RC1 to RC4 was to investigate the effect increasing the RO water recovery, thereby decreasing the volume of RO concentrate treated by the ASP-CUF system, allowing greater volumes of UF filtrate to be treated at the constant RO feed flow rate of 50 kL/d (see Figure 2.1). The purpose of run condition RC5 was to investigate the influence of RO feed flow at low RO water recovery and 90% system water recovery.

3. Results and Discussion

3.1 Major Equipment Faults during Commissioning

The major equipment faults encountered and actions taken to rectify these faults are summarised in Table 3.1.1:

Table 3.1.1: Major equipment faults encountered during commissioning

Plant Component	Equipment Fault/ problem	Potential Cause of Fault	Actions taken
DAF plant	Recirculation pump failure	Pump in poor condition, high ambient temperature, differential meter malfunction	<ul style="list-style-type: none"> • Shutdown • Replacement of pump
	Ferric chloride dosing pump failure	Pump in poor condition	<ul style="list-style-type: none"> • Shutdown • Replacement of pump
	Flocculation tank stirrer failure	Leak of lubricant	<ul style="list-style-type: none"> • Shutdown • Refilling/replacement of gear box lubricant
	Skimmer motor failure	Corrosion of skimmer chains	<ul style="list-style-type: none"> • Shutdown • lubrication of chain tracks • Refilling/replacement of gear box lubricant
	Recirculation circuit pipes burst	Differential pressure meter failure	<ul style="list-style-type: none"> • Shutdown • Replacement of pipework • Replacement of differential pressure meter
	Differential pressure (DP) meter failure	Not known	<ul style="list-style-type: none"> • Shutdown • Replacement of DP meter
UF plant	High trans-membrane pressure and low filtrate pressure	Fouling due to high water turbidity (before installation of DAF)	<ul style="list-style-type: none"> • CIP performed • Installation of DAF as pre-treatment
	Backwash tank level sensor failure (intermittent at first)	Not known	<ul style="list-style-type: none"> • Shutdown • Replacement of level sensor

Plant Component	Equipment Fault/ problem	Potential Cause of Fault	Actions taken
	" Power failure" fault warnings followed process control induced shutdown despite uninterrupted operation of generator	Not known	<ul style="list-style-type: none"> • Inspection of electrical system and generator by electrician (unable to detect a problem. This was an intermittent problem that was present throughout the trial period and more frequent towards the end of the trial period).
RO plant	Poor RO feed pump performance	Not known	<ul style="list-style-type: none"> • Pump checked by electrician • Pump sent back to manufacturer (Grundfos) to ensure programming was not faulty. • Experiments redesigned at lower feed flows and water recoveries to compensate for the poor pump performance. • (Pump replacement not an option due to cost)
	Brine conductivity meter range too low	High feedwater conductivity	<ul style="list-style-type: none"> • Replacement of conductivity meter
	Permeate conductivity meter failure	Not known	<ul style="list-style-type: none"> • Loose wire reconnected
	Burst ¼ inch pipes to pressure sensors	Piping not replaced in preparation of equipment for trial	<ul style="list-style-type: none"> • Replacement
	Severe leaks from concentrate stream (at various locations, at various times)	Loose clamps	<ul style="list-style-type: none"> • Shutdown • Replacement of seals, tightening of clamps.
	High permeate conductivity in permeate from 4 th pressure vessel	Suspected leak from seals between membrane elements	<ul style="list-style-type: none"> • Bypass of 4th pressure vessel • (Cause of probable leak of concentrate into permeate stream not investigated due to time constraints).
ASP and flocculation	Crystallisation of Na ₂ HPO ₄ in storage tank, dosing pump and in the lines.	Low ambient temperatures	<ul style="list-style-type: none"> • Shut down • Installation of heater and stirrer in the 1000L Na₂HPO₄ solution storage tank and thermal insulation of lines.
	Flash mixer stirrer failure	Not known	<ul style="list-style-type: none"> • Reinstallation of impeller

Plant Component	Equipment Fault/ problem	Potential Cause of Fault	Actions taken
CUF	High turbidity filtrate during start up	Calcium ions are able to freely diffuse through the filter and precipitate on the filtrate side at high pH.	<ul style="list-style-type: none"> • Shutdown. • Development of start-up protocol that ensures the ceramic filters are not exposed to high calcium ion concentrations.
Computers and communication	Persistent "fault" warning on RO despite rectification of fault	Not known	<ul style="list-style-type: none"> • Problem rectified by IT technician
	Intermittent communications faults	Not known	<ul style="list-style-type: none"> • Consultation with IT technician (no remedy)
General	Intermittent Power failure	Not known	<ul style="list-style-type: none"> • Consultation with electrician (no remedy)

The large number of process steps in the final plant design (Generator-DAF-UF-RO-ASP-CUF) made uninterrupted operation of the full plant for more than a few hours very difficult. Experimental results are presented as preliminary (see section 3.2) and the first longer duration run (RC2, section 3.3.1) are equivalent to commissioning results as these runs were cut short by one or more of the above faults. The results of the longer duration run RC5 (see section 3.3.2) are more than commissioning results as they give information regarding the adequacy of the plant design for this high recovery RO application. The brevity of this run, however, did not allow optimisation of chemical use and investigation of the incidence of scale formation.

3.2 Plant commissioning

The primary purpose of the preliminary runs was to test the ASP process and to see how the RO plant process control system reacts to the change in the characteristics of the RO feed water when the CUF filtrate (Ca depleted stream, see Fig 2.1) was connected to the RO. The quality of the RO feed water is shown in Table 3.2.1.

Table 3.2.1: RO feed water quality, sampled 05-05-11

Parameter	Units	Concentration
pH	pH Units	7
Phosphorus, reactive as P	mg P / L	<0.01
Total Dissolved Solids	mg/L	9500
Fluoride, as F	mg/L	0.68
Chloride, as Cl	mg/L	5600
Sulphate, as SO ₄	mg/L	970
Bicarbonate Alkalinity as CaCO ₃	mg CaCO ₃ / L	370
Carbonate Alkalinity as CaCO ₃	mg CaCO ₃ / L	<2
Hydroxide Alkalinity as CaCO ₃	mg CaCO ₃ / L	<2
Total Alkalinity as CaCO ₃	mg CaCO ₃ / L	370
Chromium 6+, as Cr	mg/L	<0.01
Chromium 3+	mg/L	<0.01
Ammonia, as N	mg N / L	11
Nitrate, as N	mg N / L	0.16
Nitrite, as N	mg N / L	0.91
Silica, reactive as SiO ₂	mg/L	15
Aluminium	mg/L	<0.1
Barium	mg/L	0.06
Boron	mg/L	1.4
Chromium	mg/L	<0.01
Copper	mg/L	0.03
Iron	mg/L	<0.2
Lead	mg/L	<0.01
Manganese	mg/L	1.7
Nickel	mg/L	0.02
Strontium	mg/L	5.3
Zinc	mg/L	0.14
Calcium	mg/L	220
Magnesium	mg/L	360
Potassium	mg/L	46
Sodium	mg/L	2800

3.2.1. ASP Process Performance During Commissioning

The calcium concentration in the CUF filtrate achieved at run conditions 1 to 4 is shown in Figures 3.2.1.1 to 3.2.1.4 (Appendix). The UF filtrate calcium concentration was approx. 220 mg/L. The expected calcium concentrations reaching the ASP system were (assuming a post-ASP calcium concentration of 10 mg/L):

- RC1: At 45% RO water recovery the feed calcium concentration to the ASP before connection to RO = 400 mg/L. After connection to RO the ASP feed calcium concentration decreased to approx 209 mg/L ($=((220 \times 25 + 10 \times 25) / 50) / 0.55$)

- RC2: At 67.5% RO water recovery the feed calcium concentration to the ASP before connection to RO = 677 mg/L. After connection to RO the ASP feed calcium concentration decreased to approx 515 mg/L ($=((220 \times 37.5 + 10 \times 12.5)/50)/0.325$)
- RC3: At 70.5% RO water recovery the feed calcium concentration to the ASP before connection to RO = 745 mg/L. After connection to RO the ASP feed calcium concentration decreased to approx 568 mg/L ($=((220 \times 37.5 + 10 \times 12.5)/50)/0.295$)
- RC4: At 73.8% RO water recovery the feed calcium concentration to the ASP before connection to RO = 840 mg/L. After connection to RO the ASP feed calcium concentration decreased to approx 695 mg/L ($=((220 \times 41 + 10 \times 9)/50)/0.262$)

The calcium removal under RC3 and RC4 were found to be better than those seen during the RC1 and RC2 experiments. This was due to better control of phosphate concentration in the Na_2HPO_4 solution used in the RC3 and RC4 experiments. The Na_2HPO_4 solution in the RC1 and RC2 experiments was less concentrated due to crystallisation of the Na_2HPO_4 during the cold nights that preceded these experiments, when the temperature fell to below 6°C. The tank was heated and continually stirred to prevent crystallisation of Na_2HPO_4 in the tank for the RC3 and RC4 experiments. It is noteworthy that the quantity of calcium entering the desalination system increased with increasing run condition number (1 to 4) as the quantity of UF filtrate increased.

3.2.2. RO Process Performance During Commissioning

The feed flow and permeate flow achieved during the preliminary runs under run conditions 1 to 4 are shown in Figures 3.2.2.1 to Figure 3.2.2.4 (Appendix). Connection and disconnection of the CUF filtrate to the RO feed line are indicated by the arrows. The RO water recovery, feed pressure and RO feed pump speed for the run condition 1 preliminary run are shown in Figure 3.2.2.5 to Figure 3.2.2.7 (Appendix).

The RO process control uses algorithms that use the following equation to drive changes in operating conditions:

$$\text{Feed flow} = (100 \times \text{Permeate flow}) / \text{Recovery (\%)}$$

The process control algorithms drives the feed pump speed to achieve the Feed flow setpoint specified by the operator, and alters the position of a brine constriction valve to achieve the pressure required to achieve the water recovery set points specified by the operator.

The return of the calcium-depleted RO concentrate to the RO (see figure 2.2.2) was expected to increase the salt concentration. The resulting increase in osmotic pressure of the feedwater was expected to decrease the permeate flow, thereby decreasing the recovery. This prompts the system to constrict the RO concentrate flow to achieve a higher pressure to drive a higher water recovery. This, in turn, prompts the pump to work harder to achieve the feed flow set point.

It can be seen in Figure 3.2.2.1 to Figure 3.2.2.4 (Appendix) that connection of the CUF filtrate to the RO feed line decreases both the feed flow and the permeate flow to varying degrees at the four different run conditions performed. This behaviour was attributed to the RO plant process control attempting to meet the water recovery setpoint under conditions of increasing feed ionic strength

(increasing osmotic pressure), by constricting the brine stream flow in order to generate a higher pressure (see Figure 3.2.2.5 and Figure 3.2.2.6). This, in turn, decreases the feed flow and drives the RO feed pump to higher speed to maintain the feed flow set point (see Figure 3.2.2.7, Appendix). Once the pump reached maximum speed, however, it could not maintain the feed flow set conditions (compare Figure 3.2.2.1 to Figures 3.2.2.5, Figure 3.2.2.6 and Figure 3.2.2.7, Appendix.) The RO feed pump's inability to meet the feed flow set point in the latter stages of the operation at RC1 to RC4 process conditions, prompted the inclusion of RC5 in the experimental plan (see Table 3.2.1). Operation at a lower feed flow setpoint decreased the demand on the pump, allowing operation for long periods without significant deviation from the flow set points (see section 3.3.2).

3.3 Experimental trials

Two longer duration trials were performed – one at run condition 2 (RC2) and one at run condition 5 (RC5, see Table 2.1).

The primary objectives of the RC2 run were:

- To repeat the previous RC2 run, which was cut short by a severe leak of the concentrate lines;
- To determine the approximate minimum phosphate requirement for good calcium removal, for further optimization in longer duration runs.

The primary objective of the RC5 run was to run for long enough, at constant feed and permeate flow, to allow further optimisation of the phosphate addition and evaluation of the incidence of scale formation

3.3.1 RC2 Experiment

3.3.1.1 ASP performance and phosphate dosing requirements

Optimisation of phosphate addition

The ASP equipment was started on 260 mL/min of saturated 30° C calcium phosphate solution, delivering an excess of phosphate (approximately 2 times the calcium input, at 220 mg/L Ca, at 50 kL/d RO feed flow). Within the first 8 hours, the calcium concentration of the CUF filtrate was found to increase and then decrease in a manner consistent with previous runs (see Figures 3.2.1.1 to 3.2.1.4). In the next 2 hours, however, the calcium concentration was found to increase from approx. approximately 25 mg/L to approximately 35 mg/L and remain at 35 mg/L for the next 7 hours. At 9:30 hours, the CUF filtrate line was connected to the RO feed line, and the calcium concentration was found to decrease to below 10 mg/L within the next two hours. The improvement in calcium removal may, at least partly, be due to the decrease in UF filtrate flow (see Figure 2.1.1) that results when some of the UF filtrate is replaced by calcium-depleted CUF filtrate at constant RO feed flow, bringing in less calcium into the ASP process. The phosphate dosing rate was decreased to 210 mL/min at 12:30 hours. The calcium concentration remained low for the next 2 hours at this lower phosphate dosing rate. The CUF filtrate line was disconnected from the RO feed line at 14:00 hours due to a severe leak in the RO concentrate line due to the high pressure (see Figure 3.3.1.2.6:and compare to Figure 3.2.2.6), necessitating shutdown of the RO. The leak was fixed and the RO was turned back on. The RO concentrate line to the ASP was disconnected at 14:00 hours and

reconnected at 15:00 hours. The phosphate dosing rate was further decreased to 152 mL/min at 15:00 hours, after which the calcium concentration in the CUF filtrate was found to increase to 60 mg/L.

The minimum phosphate requirement for this set of conditions was found to be between 152 mL/min and 210 mL/min. Using the 210 mL/min figure:

- The quantity of disodium hydrogen phosphate delivered per minute:

$$= 0.210 \text{ L/min} \times 200 \text{ g/L}$$

$$= 42 \text{ g/min}$$

$$= 42/142 \text{ mole/min}$$

$$= 0.296 \text{ mole/min}$$

- The quantity of calcium delivered to the ASP plant per minute (as per RC2):

$$= \text{UF filtrate feed flow} \times \text{UF filtrate feed Ca conc.}$$

$$= (37500/24/60 \text{ L/min}) \times 0.220 \text{ g/L}$$

$$= 5.7 \text{ g/min}$$

$$= 5.7/40 \text{ mole/min}$$

$$= 0.142 \text{ mole/min}$$

$$\text{Ratio of } \text{PO}_4^{3-} \text{ to } \text{Ca}^{2+} = 0.296/0.142 = 2.1$$

It is noteworthy that the laboratory experiments indicated that the quantity of phosphate required was expected to decrease to less than equimolar to the calcium input as the quantity of in-situ generated seed builds up in the ASP tank with time as a result of solids recycling. Build up of the seed concentration to useful levels (10 to 20 g/L) was, however, prevented by the short run duration (5 hours). The time required for seed concentrations to increase to useful levels is discussed in Section 3.3.2.1.1 (p. 31).

Turbidity of CUF filtrate

The turbidity results for Run Condition 2 are shown in Table 3.3.1.1.1 (see Appendix, p.62). It can be seen from the results that the ceramic filter consistently achieved turbidities well below 1 NTU up to 20:00 hours of Day 2, after which the turbidity rapidly increased to above the meter's limit (1,000 NTU). The generation of a turbid ceramic filter filtrate was also found to occur during experiment RC5. See Section 3.3.2.1, on p.30 for further discussion regarding the cause of the increase in turbidity of CUF filtrate.

It is interesting to note that the ceramic filtration system imparts a considerable temperature increase to the process streams. The rapid increase in temperature between 14:00 and 15:00 hours seen in Figure 3.3.1.1.4 was the result of recirculating (with no filtration and no waste flow) of the seed within the filtration equipment to prevent sediment cake formation while the RO leak was being repaired. This temperature increase occurred after a slower increase from 30 to 35° C which occurred after connection of the CUF filtrate to the RO at 9:30 hours. The RO feed temperature was approximately 12° C before connection of the CUF filtrate to the RO.

3.3.1.2 RO performance

It can be seen that connection of the CUF filtrate to the RO feed line decreased both the feed flow and the permeate flow (see Figure 3.3.1.2.2). A similar outcome was seen in the preliminary runs (see section 3.2). This behaviour can be attributed to the RO plant process control's inability to meet the water recovery set point under conditions of increasing feed ionic strength (increasing osmotic pressure, Figure 3.3.1.2.10) due to the pump having already reached its upper limit of speed (see Figure 3.3.1.2.8) before connection of the CUF filtrate to the RO feed line.

3.3.2 RC5 Experiment

All runs (RC1 – RC4) prior to this run were characterised by a decrease in feed flow and permeate flow on connection of the CUF filtrate to the RO feed line. This behaviour was attributed to the RO plant process control's inability to meet the water recovery set point under conditions of increasing feed ionic strength. This was due to the pump having already reached its upper limit of speed before connecting the CUF filtrate to the RO feed line. This run attempted to overcome this situation by running at lower feed flow set point and at low RO water recovery (see Section 3.2.2).

The target flows for this run are summarised in Figure 3.3.2.1. However, an abrupt change in flow reading from the CUF filtrate flow meter at 4:30 hours on Day 2 (see Figure 3.3.2.1.8) indicates that the actual CUF filtrate flow during the first 20 hours of this run was approximately 500 L/h, giving a waste flow of approx. 300 L/h and a system water recovery (with constant feed flow to RO) of approx. 70% during this time period. The waste stream flow was timed and found to be approx. 300 L/h.

Adjustment of the CUF filtrate flow to approx 700 L/h resulted in immediate increase in the CUF filtrate turbidity (see Figure 3.3.2.1.5), necessitating the disconnection of the CUF filtrate from the RO feed line to protect the RO. The turbidity of the CUF filtrate stream was found to clear after 8 hours of continued operation of the ASP (see Figure 3.3.2.1.6), with the CUF filtrate flow going to waste. The CUF filtrate stream was reconnected to the RO at 17:00 hours on Day 2, allowing operation until 2:00 hours of Day 3, at which point another turbidity increase in the CUF filtrate stream was recorded (see Figure 3.3.2.1.7). The average system waste flow between 17:00 hours of Day 2 and 2:00 hours of Day 3 was 82 L/h, equating to a 90% desalination system water recovery during this period.

The CUF filtrate stream was disconnected from the RO feed stream at approx. 2:00 hours on Day 3 and continued operation of the ASP-CUF system for a further 10 hours was not found to improve the turbidity. The run was stopped at 12:30 hours on Day 3.

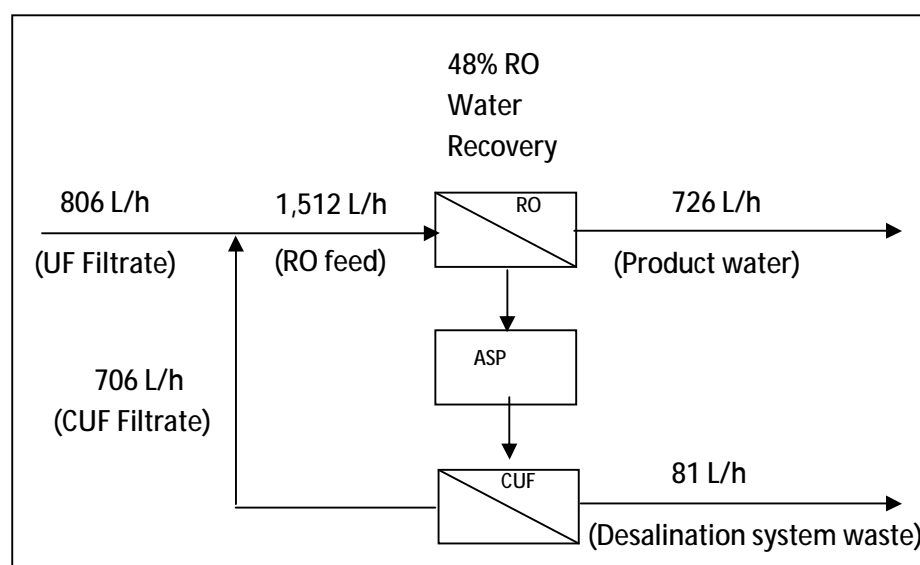


Figure 3.3.2.1: RC5 target flows

3.3.2.1 ASP performance and phosphate dosing requirements

The RC5 operating conditions and results are discussed in Table 3.3.2.1.1. The results are presented in Figures 3.3.2.1.1 to Figure 3.3.2.1.14.

Table 3.3.2.1.1: Discussion of RC5 operating conditions and results

Day and Time	Reference Data	Discussion
Day 1, 13:00 to Day 1, 17:00	Figure 3.3.2.1.2	<p>The UF filtrate calcium concentration on RC5, Day1 was 198 mg/L. At 48% RO water recovery, the ASP feed calcium concentration before connection to RO was 412 mg/L. This was reduced to less than 20 mg/L by the ASP process. This equates to 95% calcium removal.</p> <p>The phosphate dosing rate at the beginning of this run was 200 mL/min of 200 g/L Na₂HPO₄ solution. At this phosphate dosing rate the number of mole added per hour = 16.9 mole /h (= (0.2 x 200 x 60)/142).</p> <p>Without connection of CUF filtrate to RO (feed flow = 1512 L/h, see Figure 3.3.2.1, no CUF filtrate) containing 198 mg/L Ca, the calcium entering the system = 7.5 mole/h (= (0.198 x 1512)/40).</p> <p>The HPO₄ to Ca mole ratio was, therefore, 16.9: 7.5 = 2.2 during this start-up period.</p> <p>The calcium concentration entering the ASP at 48% RO water recovery was 380 mg/L. The calcium concentration immediately before connection was 20 mg/L, equating to a removal of 95%.</p>
Day 1, 17:00 to Day 2, 7:00 ₃	Figure 3.3.2.1.2 and Figure 3.3.2.1.	<p>After connection to the RO, the blending of the calcium depleted stream with a smaller volume of the UF filtrate stream decreases the calcium concentration reaching the ASP to 212 mg/L (= ((806 x 198 + 706 x 10)/1512)/0.52, assuming the post-ASP calcium concentration is 10 mg/L) = 5.3 mmole/L. This was reduced to less than 10 mg/L (0.25 mmole/L) soon after connection to the RO (see Figure 3.3.2.1.2).</p> <p>Connection of CUF filtrate to RO decreased the quantity of raw water entering the system and the calcium entering the system to 3.8 mole/h (= 0.198 x 806)/40).</p> <p>Subsequent reduction of Na₂HPO₄ dose to 156 mL/min decreased the number of mole of phosphate added to 13.2 mole/h (= (0.156x60x200)/142), HPO₄ to Ca mole ratio = 13.2 : 3.8 = 3.5) and increased the calcium level in the CUF filtrate to near the pre-connection level of 20 mg/L.</p> <p>Further optimisation of phosphate dose was postponed due to concerns regarding the possibility of generation of insoluble calcium salts (phosphate, carbonate, hydroxide) in the filtrate side of the ceramic filters (as was found to occur in RC2 run, see section 3.3.1.1) from the available soluble calcium at elevated temperatures (28 Deg. C, see Figure 3.3.2.1.11).</p> <p>The phosphate concentration was increased to 180 mL/min (15.2 mole/h, HPO₄ to Ca mole ratio = 4.0) for the remainder of this period to keep the calcium concentration low.</p> <p>It is important to note that prolonged periods of operation at high system water recoveries with large excesses of phosphate such as these are conducive to calcium phosphate formation on the membranes, and such high phosphate additions cannot be sustained over long periods. It was, however, deemed more important at this stage of the trial to avoid a turbidity increase in the CUF filtrate.</p>

Day and Time	Reference Data	Discussion
Day 2, 7:30 to Day 2, 17:00	Figure 3.3.2.1.3 and Figure 3.3.2.1.6	<p>Concern regarding the possibility of generation of insoluble calcium salts on the filtrate of the ceramic filter appears to have been justified. An increase in filtrate flow (see section 3.3.2) was found to give rise to very turbid ceramic filter filtrate (see Figure 3.3.2.1.6). Disconnection of the CUF filtrate from the RO feed line was performed at 7:30 hours to protect the RO.</p> <p>The calcium concentration after disconnection of the ASP is expected to increase from 212 back up to 380 mg/L. The phosphate feed flow was increased to 200 mL/min (16.8 mole/h, HPO₄ to Ca mole ratio = 4.4) at 7:30 hours, and to 240 mL/min (20.4 mole/h, HPO₄ to Ca mole ratio = 5.4) at 10:00 hours to ensure rapid decrease of the calcium concentration.</p> <p>Continued filtration at low calcium concentration was found to flush the precipitated insoluble calcium species in the ceramic filter filtrate line, eventually giving a turbidity less than 1 NTU (see Figure 3.3.2.1.6).</p>
Day 2, 17:00 to Day 3, 2:30	Figure 3.3.2.1.3, Figure 3.3.2.1.4 and Figure 3.3.2.1.7	<p>The CUF filtrate was reconnected to the RO feed line at 17:00 hours on Day 2, and the plant was operated at a phosphate dose rate of 180 mL/min (0.25 mole/min) until 2:00 hours on Day 3, when another increase in filtrate turbidity was observed. The CUF filtrate was, once again, disconnected from the RO feed line.</p> <p>It is important to note here that the calcium concentration had remained consistently low (less than ~8 ppm) for 9 hours before the turbidity spike occurred, and that further phosphate reductions may have been possible without greatly increasing the calcium concentration had the turbidity spike not occurred. Furthermore, laboratory trials early in the project indicated that the increase in seed loading improves removal, necessitating less phosphate addition. It is expected that the phosphate requirement would decrease as the amount of seed increased in the system (see discussion: "Effect of Seed Recycling, p.31).</p>
Day 3 2:30 to Day 3, 12:30	Figure 3.3.2.1.3, Figure 3.3.2.1.7	<p>Attempts to flush out the precipitated solids as previously performed on day 2 proved ineffective. During the subsequent 10 hours of plant operation the ceramic filter filtrate turbidity did not decrease below 1000 NTU (the turbidity meter limit). The run was stopped at 12:30 hours.</p>

Turbidity of CUF filtrate

Two increases in CUF filtrate turbidity were found to occur during this run. One at the end of the 70% system water recovery period (Day 1, 17:00 to Day 2, 7:00) and one at the end of the 90% system water recovery period (Day 2, 17:00 to Day 3, 2:30). During these periods, the temperature of the CUF filtrate was found to increase (see Figure 3.3.2.1.11, Figure 3.3.2.1.12 and Figure 3.3.2.1.13). The higher temperature is the result of the high internal filter flow rates used to minimise the chance of cake formation (70,000 L/h), and due to the increase in temperature of the RO feed resulting from the blending of the warm CUF filtrate and the cooler (~12 C) UF filtrate.

An increase in temperature is known to enhance precipitation of calcium phosphate due to the lower solubility of these salts at the higher temperature. The increase in CUF filtrate turbidity can then, tentatively, be attributed to precipitation of calcium species on the filtrate/permeate side of the ceramic filter when the calcium and the anionic insoluble calcium salt precursor ions (e.g. CO_3^{2-} , HPO_4^{2-}) are exposed to higher temperatures in this region of the filtration equipment. Since the increase in temperature is imparted to the retentate/concentrate stream by the pump, this mechanism of solids formation on the filtrate/permeate side of the ceramic filter would require an induction period that is sufficiently long as to allow the free ion to travel to the filtrate side before precipitation takes place (see Figure 3.3.2.1.1). Further testing is required to test this speculative explanation. If the cause is the increase in temperature, the use of a heat exchanger, lower cross flow velocities and/or higher feed flows may minimise this effect and allow longer periods of operation using ceramic filters.

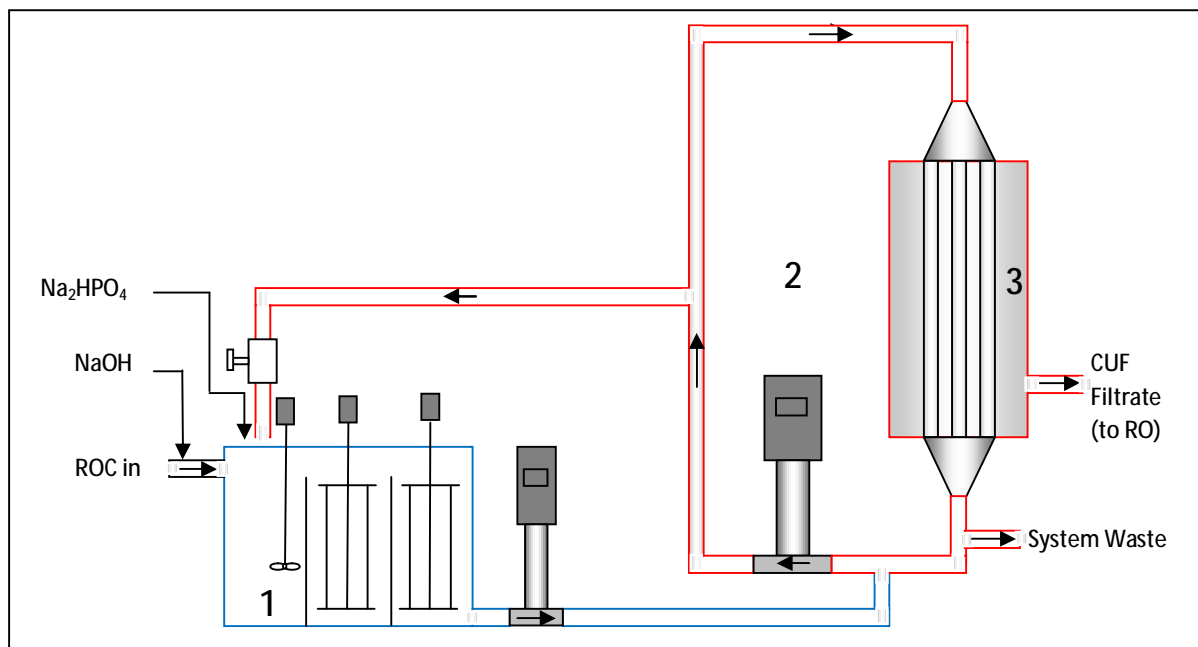


Figure 3.3.2.1.1: Diagrammatic representation of the ASP-CUF system and proposed mechanism for generation of turbidity in the CUF filtrate line; 1: Rapid precipitation of CaHPO_4 to the solubility limit in the flash mixer and flocculation tanks; 2: Heating due to pump energy input in the ceramic filter recirculation loop (flow = 70,000 L/h) and generation of a supersaturated calcium solution due to decreased solubility at the higher-temperature; 3: slow precipitation of sparingly soluble calcium compounds to the higher-temperature solubility limit in the filtrate.

3.3.2.1.1 Effect of seed recycling

The lower system water recycling in the first 8 hours of this run (see paragraph 2 in Section 3.3.2) offers the opportunity to observe the effect of solids recycling:

At 70% system water recovery: With a UF filtrate calcium concentration of 198 mg/L, the quantity of calcium entering the system at 1026 was $0.198 \times 806 = 159$ g/h as calcium ion which equates to 690 g/h of CaHPO_4 . This weight of CaHPO_4 would be expected to be confined to the ASP flocculation tanks (500 L) and the ceramic filter system (100L). In the first hour, the loss of solids expected from the system in the waste stream (300 L/h for the first 8 hours of operation) would be $690 \times (300/600) = 345$ g, leaving 345 g of CaHPO_4 with a concentration of $345/600$ g/L = 0.57 g/l in the first hour of operation. Application of this calculation (690 g/h of CaHPO_4 entering the system per hour with 50% being lost in waste per hour) to subsequent hours reveals that the maximum calcium phosphate seed concentration achieved was 0.9 g/L after approx 5 hours (see Figure 3.3.2.1.13)

At 90% system water recovery: Similarly, in the first hour, the loss of solids expected from the system in the waste stream at 90% system water recovery (82 L/h) would be $542 \times (82/600) = 74$ g, leaving 568 g of CaHPO_4 with a concentration of 0.78 g/L ($568/600$) in the first hour of operation. Application of this calculation (542 g/h of CaHPO_4 entering the system per hour with 13.7 % being lost per hour in the waste) to subsequent hours reveals that the maximum calcium phosphate seed concentration achieved was approx. 5.7 g/L after approx 30 hours (see Figure 3.3.2.1.14). After 8 hours of operation, the expected solids concentration was approximately 4.5 g/L.

The higher calcium removal achieved at the end of the 90% water recovery period (Day 2 18:00 to Day 3 2:00) compared to the removal at the end of the 70% water recovery period (Day 1 17:00 to Day 2 7:00) can be attributed to the higher solids content of the former (4.5 g/L) than the latter (0.9 g/L). Both performance conditions were at similar phosphate dose rates (180 mL/min).

It is noteworthy that laboratory experiments have shown that solids concentrations of approximately 17 g/L are required to reduce the phosphate requirement to approximately one half to one third of equimolar to the incoming calcium level (see Figure 2.8 of Milestone 3 report). This seed concentration could be achieved by taking more than the required quantity of CUF filtrate and diverting some to waste. The ceramic filtration unit was set up to be able to do this and operation in high solids mode was planned for the latter stages of the run. The premature end of the run due to high CUF filtrate turbidity prevented this investigation, and further optimisation of the phosphate addition rate.

Fouling of Ceramic Filters

Only slight adjustment of the CUF filtrate flow valves was necessary during the trial, indicating that the fouling tendency was low for the ceramic filter at the high recirculation flow and relatively low solids loading (up to ~6 g/L after 24 hours) operating conditions.

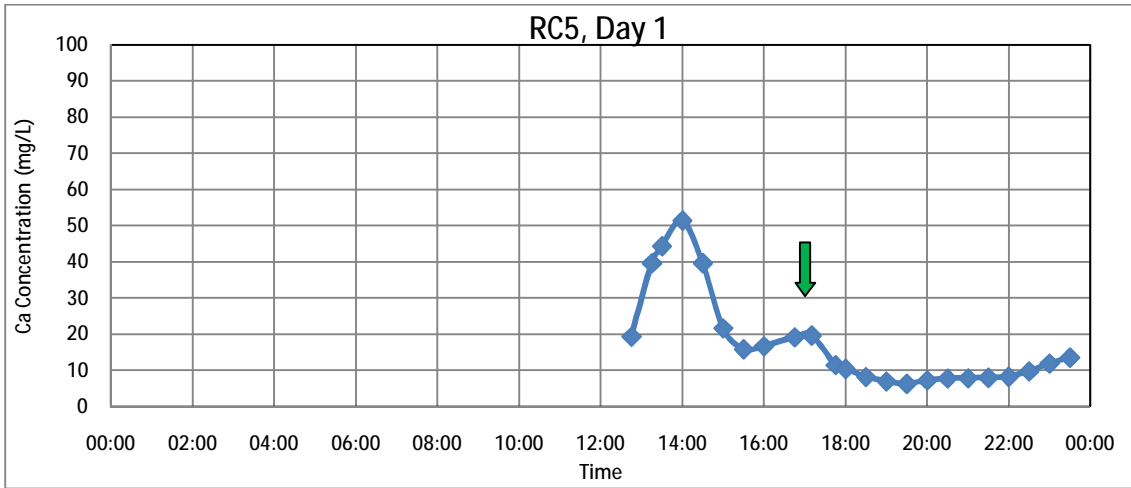


Figure 3.3.2.1.2: CUF filtrate calcium concentration, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow, initial Ca conc. = 380 mg/L

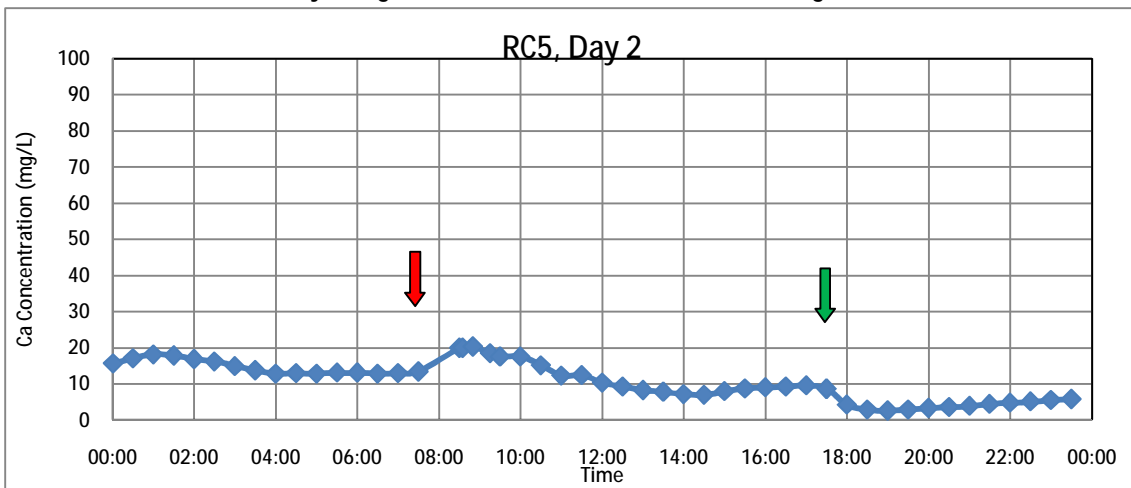


Figure 3.3.2.1.3: CUF filtrate calcium concentration, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by green and red arrows respectively, initial Ca conc. = 380 mg/L

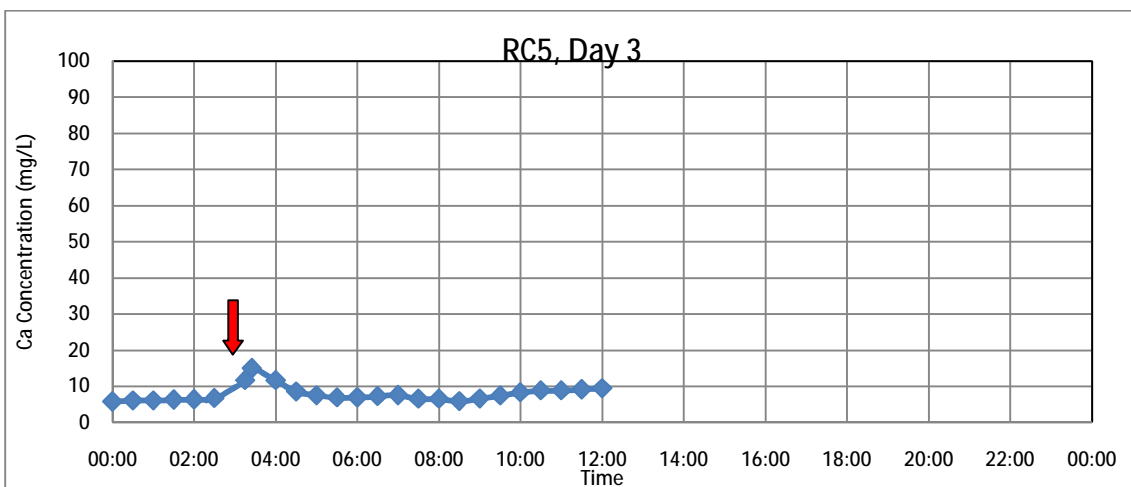


Figure 3.3.2.1.4: CUF filtrate calcium concentration, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow, initial Ca conc. = 380 mg/L,

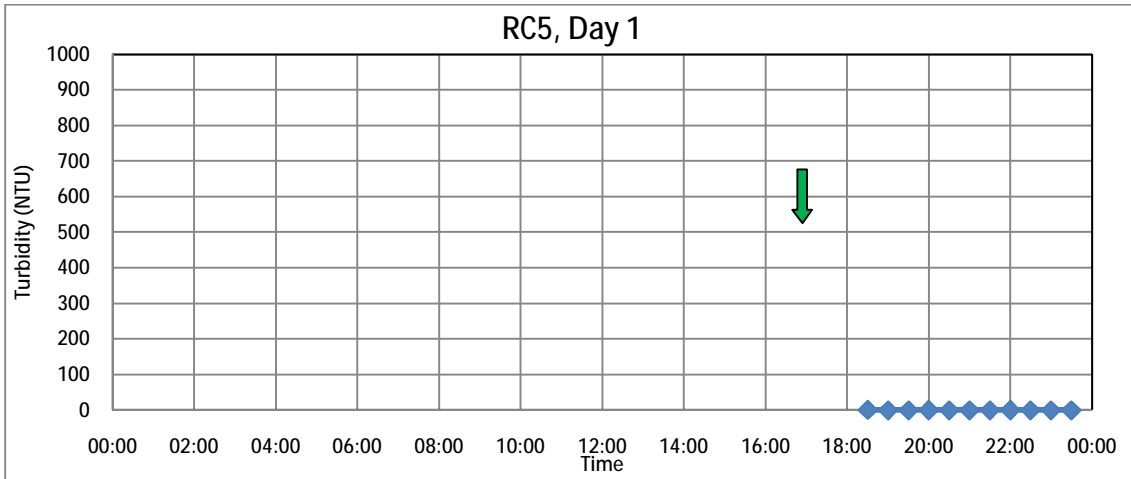


Figure 3.3.2.1.5: CUF filtrate turbidity, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by green arrows

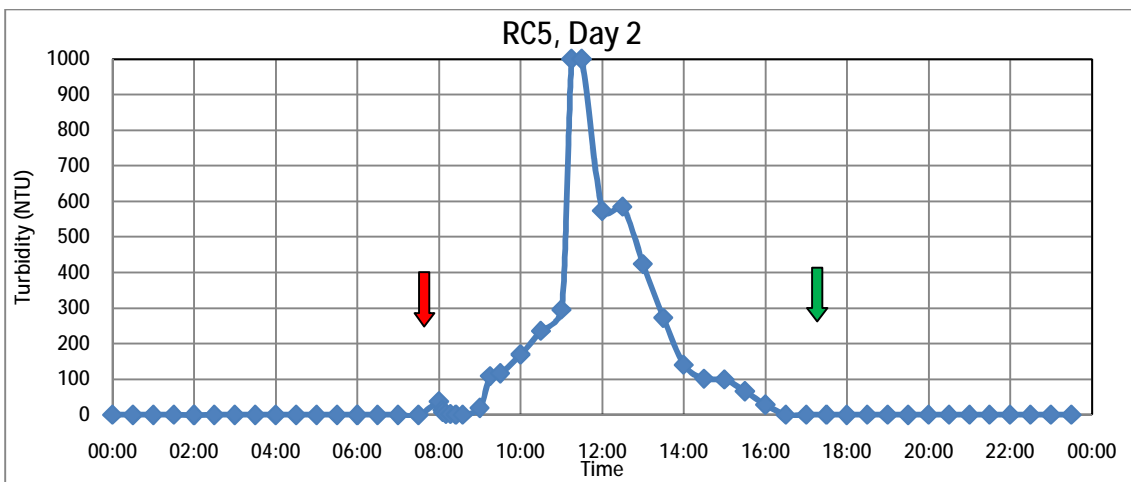


Figure 3.3.2.1.6: CUF filtrate turbidity, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by green and red arrows respectively

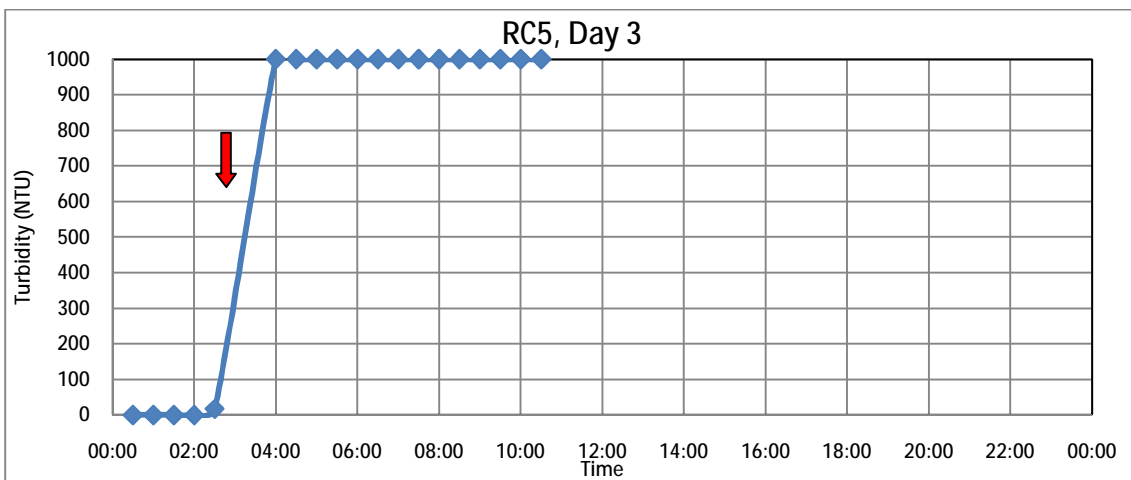


Figure 3.3.2.1.7: CUF filtrate turbidity, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by red arrow

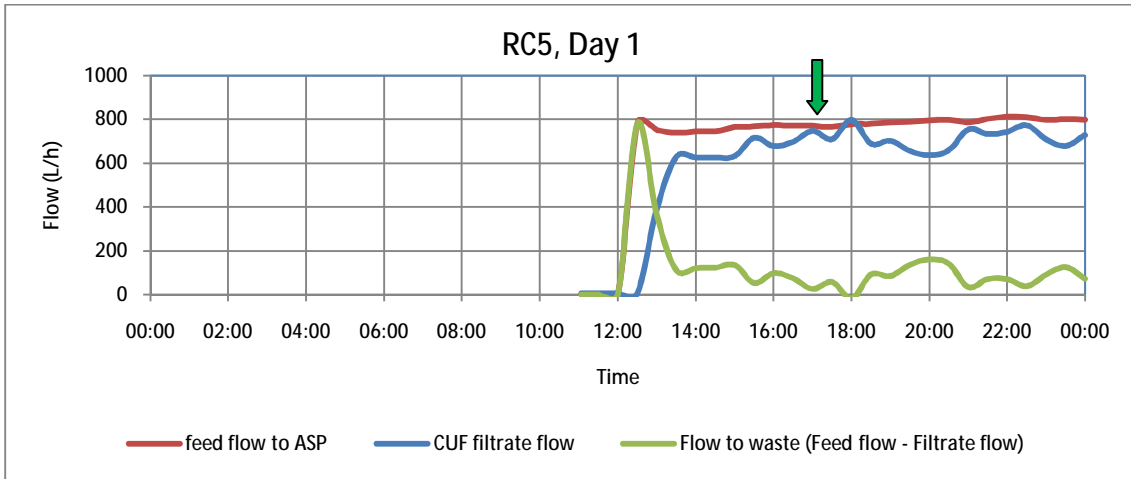


Figure 3.3.2.1.8: ASP stream flows, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrows

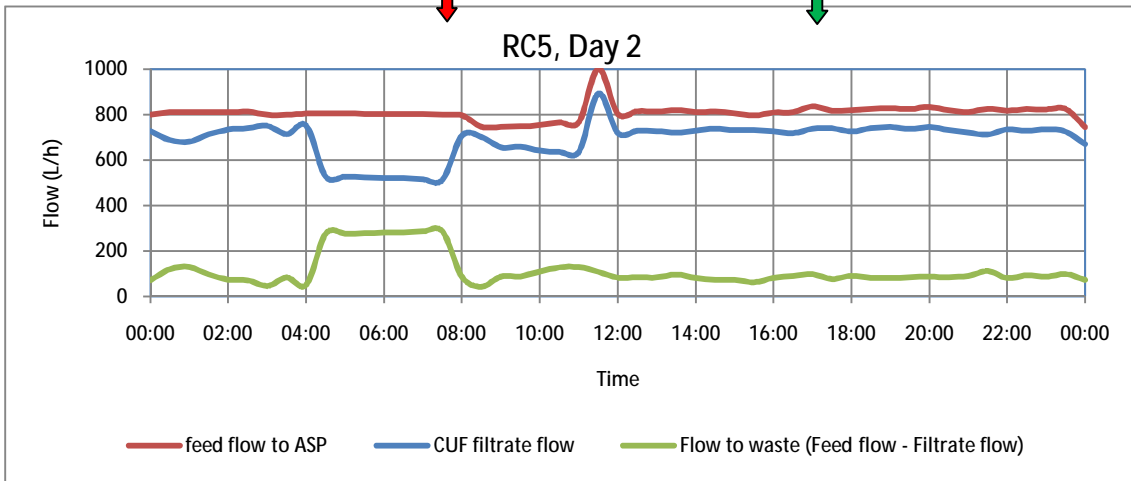


Figure 3.3.2.1.9: ASP stream flows, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

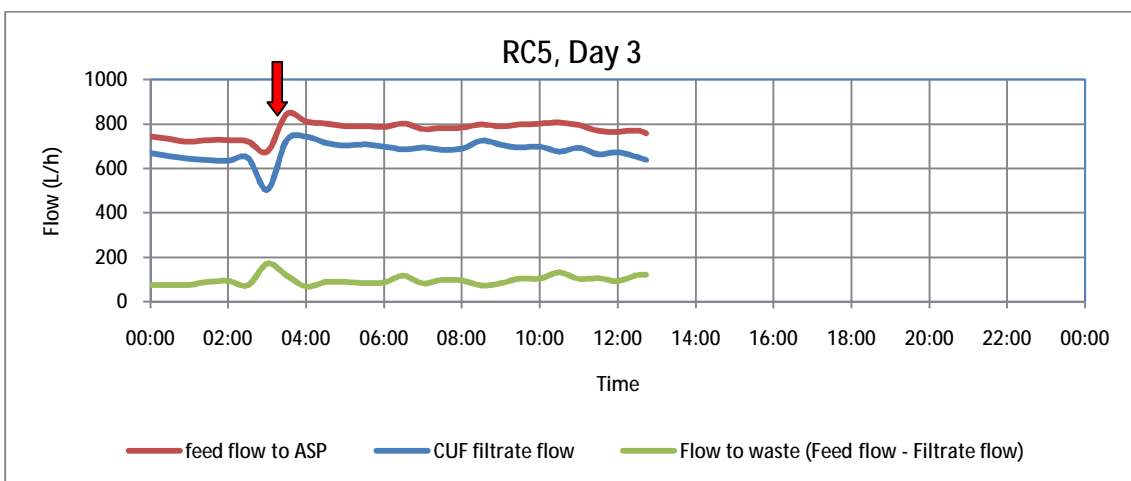


Figure 3.3.2.1.10: ASP stream flows, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

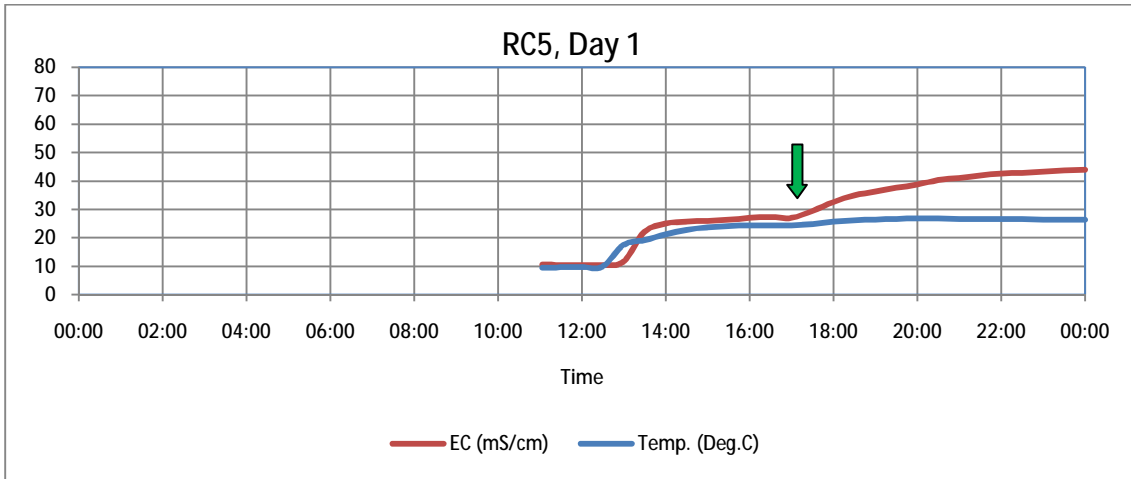


Figure 3.3.2.1.11: ASP stream electrical conductivity (EC) and temperature, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

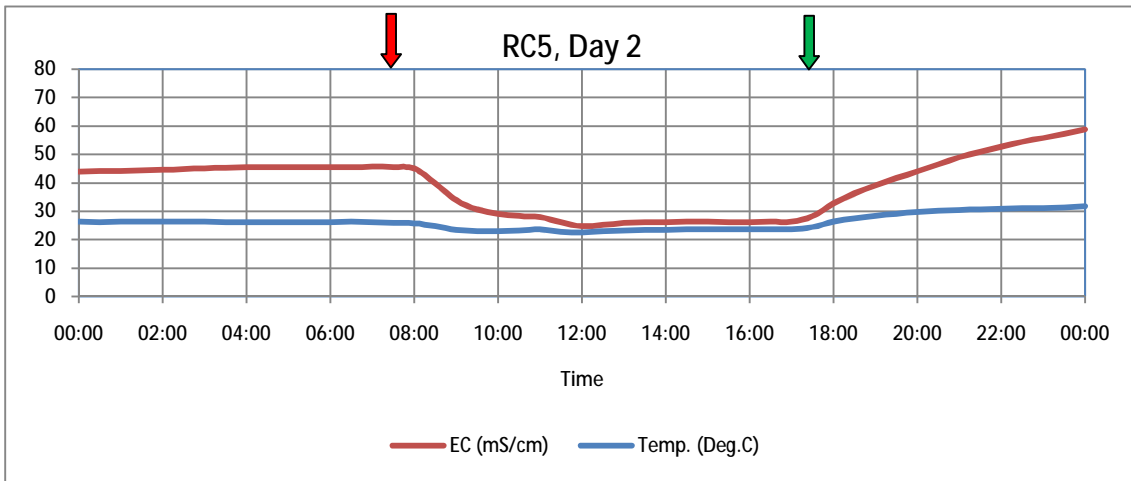


Figure 3.3.2.1.12: ASP stream electrical conductivity (EC) and temperature, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

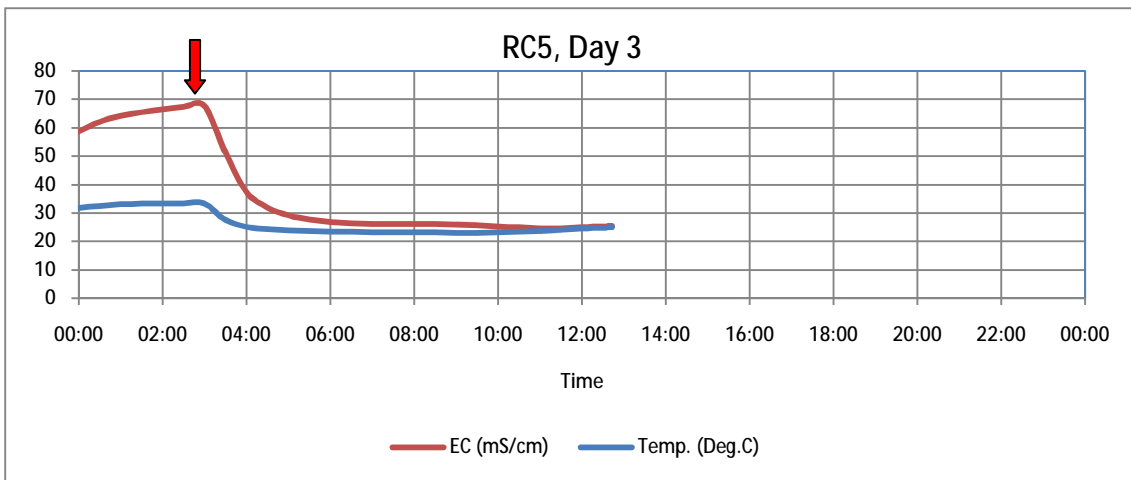


Figure 3.3.2.1.13: ASP stream electrical conductivity (EC) and temperature , RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

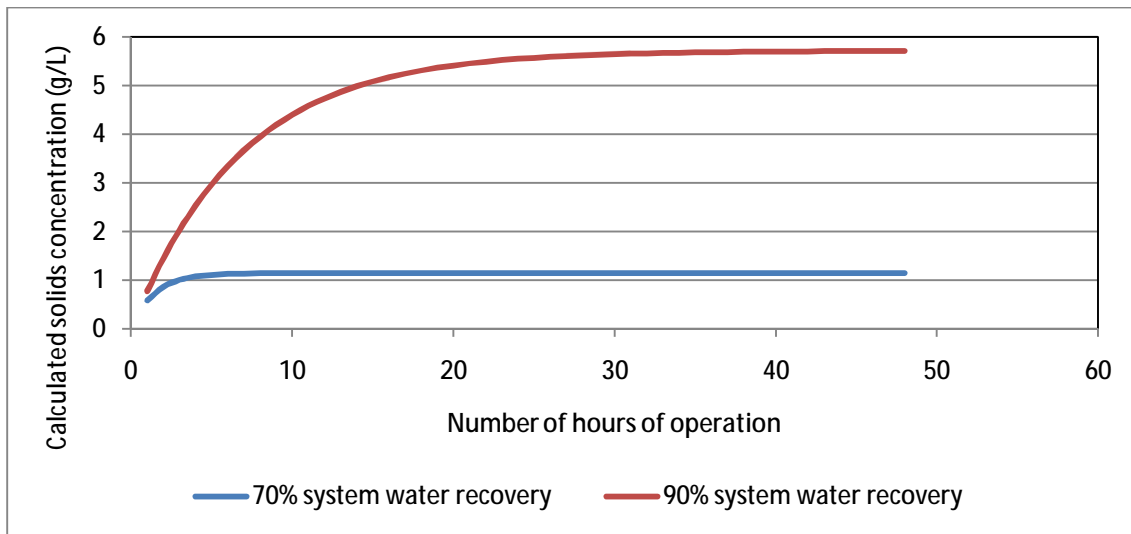


Figure 3.3.2.1.14: Calculated seed concentration in ASP-CUF system at the achieved system water recoveries, RC5.

3.3.2.2 RO performance

The decrease in feed and permeate flow upon connection of the CUF filtrate to the RO feed line, which was characteristic of previous runs at higher RO feed flow set points, was not observed during RC5 (see Figure 3.3.2.2.1 to Figure 3.3.2.2.3 and compare to Figure 3.3.1.2.1 and Figure 3.3.1.2.2).

The RO recovery remained steady at the set point until disconnection of the CUF filtrate stream from the RO feed stream at 8:00 hours on Day 2 (see Figure 3.3.2.2.4 and Figure 3.3.2.2.5). The sudden disconnection led to a small increase in RO water recovery due to the lower feed salt content (see Figure 3.3.2.2.14), prompting the brine constriction valve to open until it reached the 70% open position (see Figure 3.3.2.2.5), at which point the RO feed pump speed (see Figure 3.3.2.2.10) was decreased and then increased again by the process control to achieve the RO water recovery set point of 48%.

Reconnection of the CUF filtrate to the RO feed line at 17:00 hours on Day 2 slightly decreased the RO water recovery below the set point due to the higher salt content of the feed (see Figure 3.3.2.2.14), prompting the brine constriction valve to close (see Figure 3.3.2.2.4) and the pump to work harder to generate a higher driving pressure (see Figure 3.3.2.2.8).

Disconnection of the CUF filtrate stream from the RO feed stream at 3:00 hours on Day 3 resulted in the return of RO operating conditions to similar values to those at the beginning of the experiment. The decrease in specific flux during the periods when the CUF filtrate was connected to the RO (17:00 Day 1 to 9:30 hours Day 2, 17:00 Day 2 to 3:00 hours Day 3) can be attributed to the increase in ionic strength and temperature during these periods. Comparison of the specific flux (Figure 3.3.2.2.25, Figure 3.3.2.2.26, Figure 3.3.2.2.27) for the constant ionic strength and temperature periods during which the CUF filtrate was disconnected from the RO feed (12:00 to 15:00 Day 1, 7:00

to 15:00 Day 2, 3:00 to 12:00 Day 3) reveals a gradual decrease in specific flux, from 1.8 LMH/Bar on Day 1, to 1.7 LMH/Bar on Day 2, to 1.65 LMH/Bar on Day 3. This represents an 8% decrease in specific flux over the three days. This scaling may be attributable to the formation of calcium phosphate on the membranes due to the excess of phosphate leaving the ASP stage. Dow Filmtec recommends that a drop of 10% in the normalised specific flux is an indication of fouling.

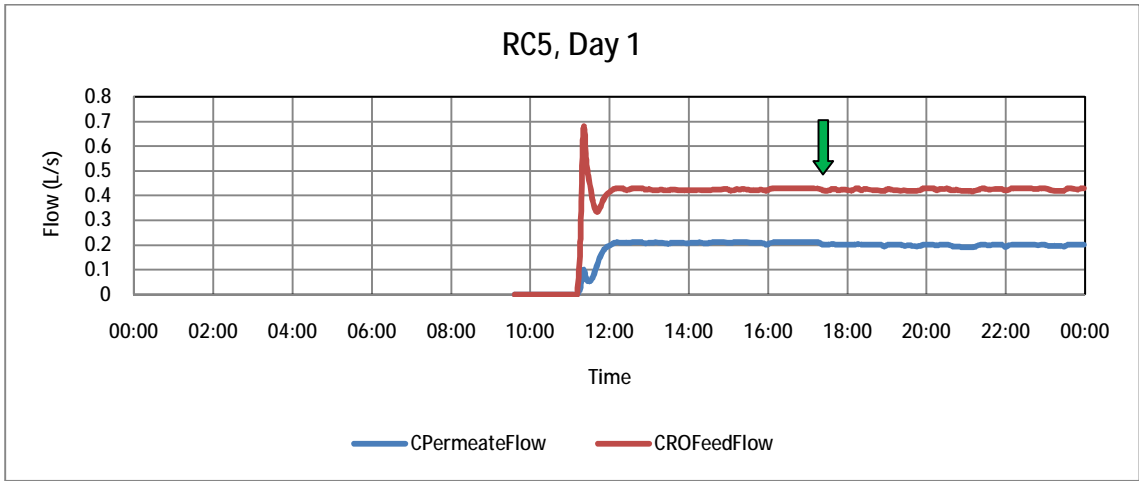


Figure 3.3.2.2.1: RO Feed and permeate flows, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by green arrow

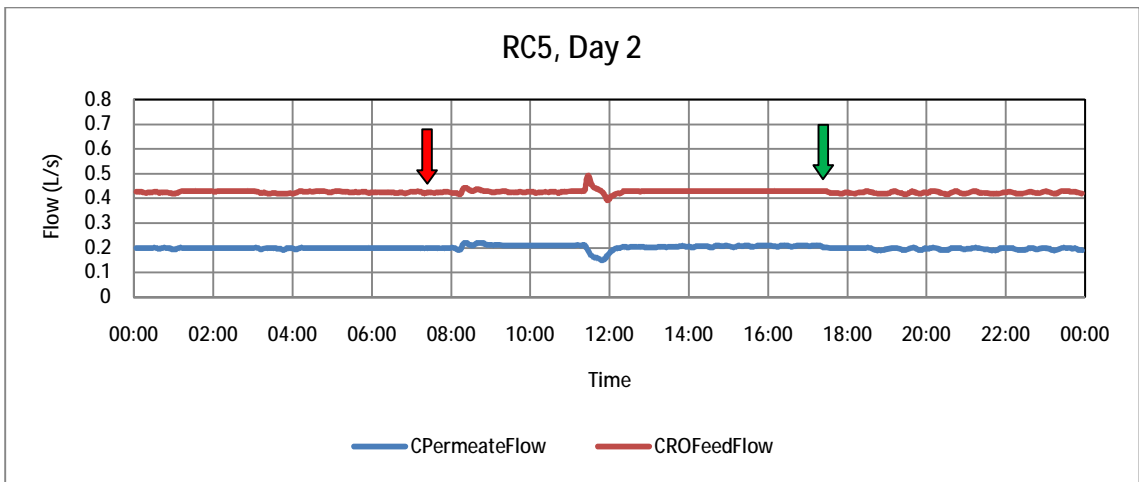


Figure 3.3.2.2.2: RO Feed and permeate flows, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

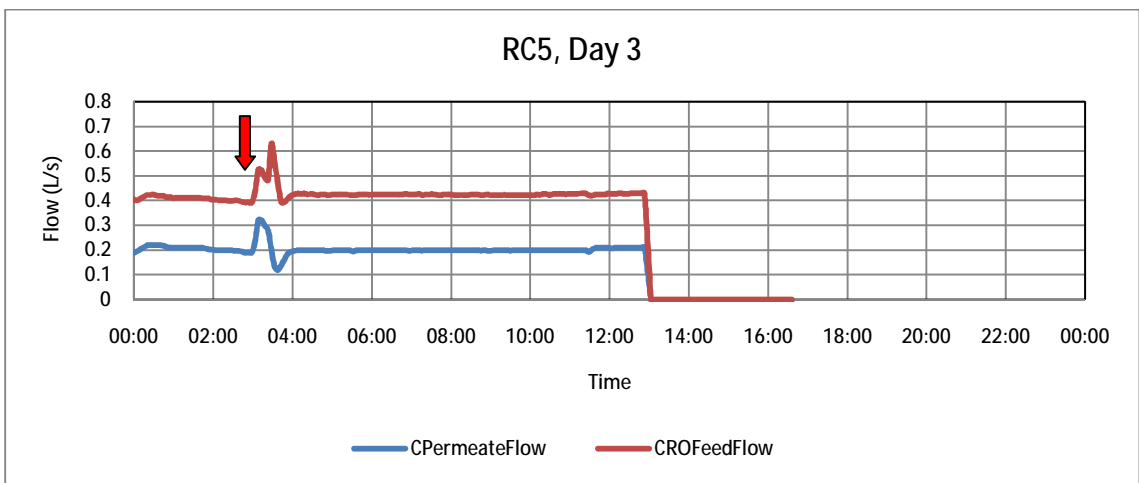


Figure 3.3.2.2.3: RO Feed and permeate flows, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

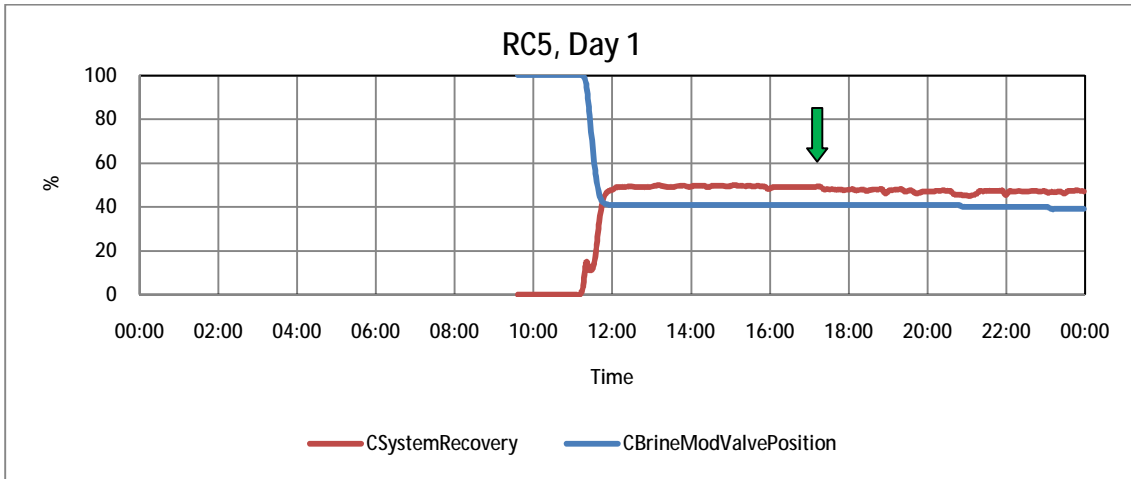


Figure 3.3.2.2.4: RO recovery and brine constriction valve position, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

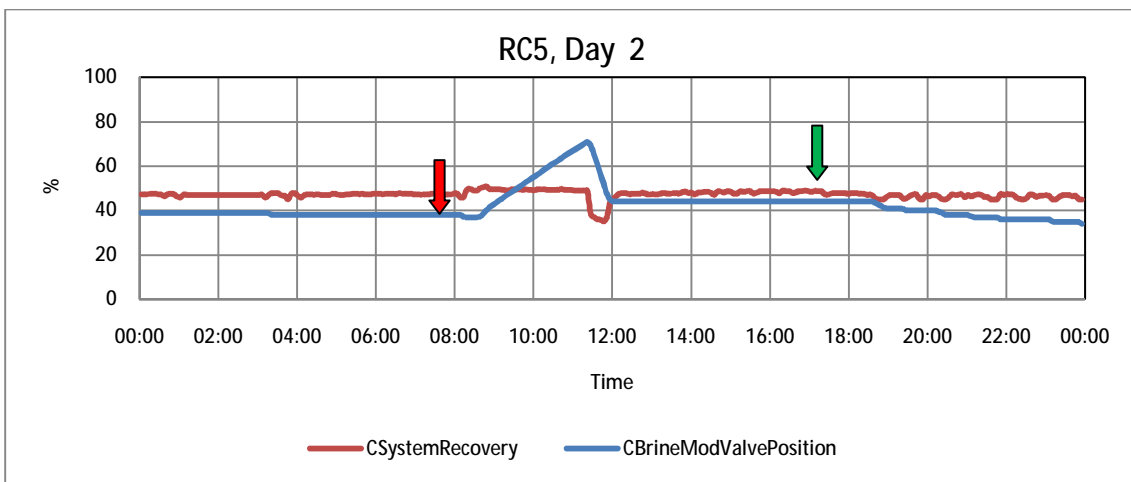


Figure 3.3.2.2.5: RO recovery and brine constriction valve position, RC5, Day 2 connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

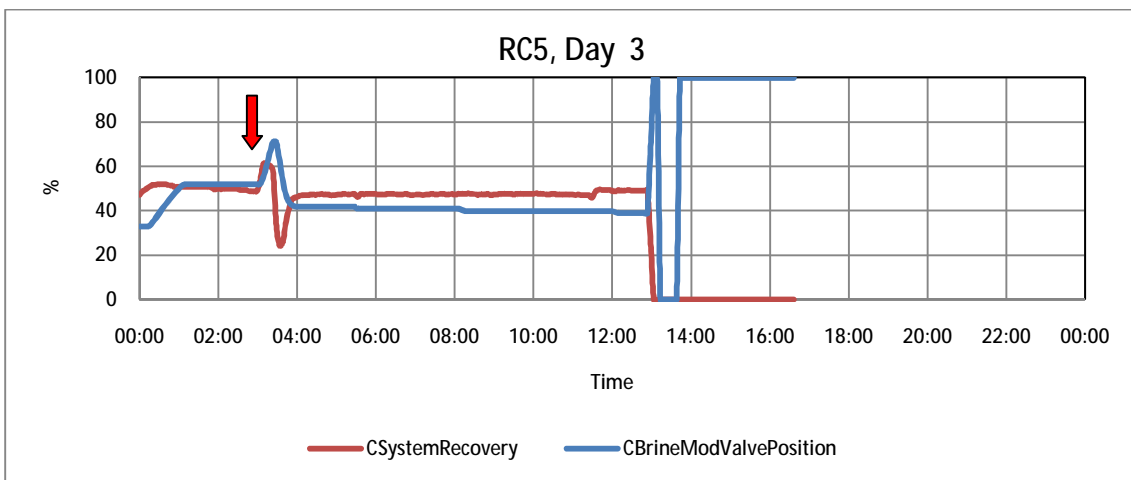


Figure 3.3.2.2.6: RO recovery and brine constriction valve position, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

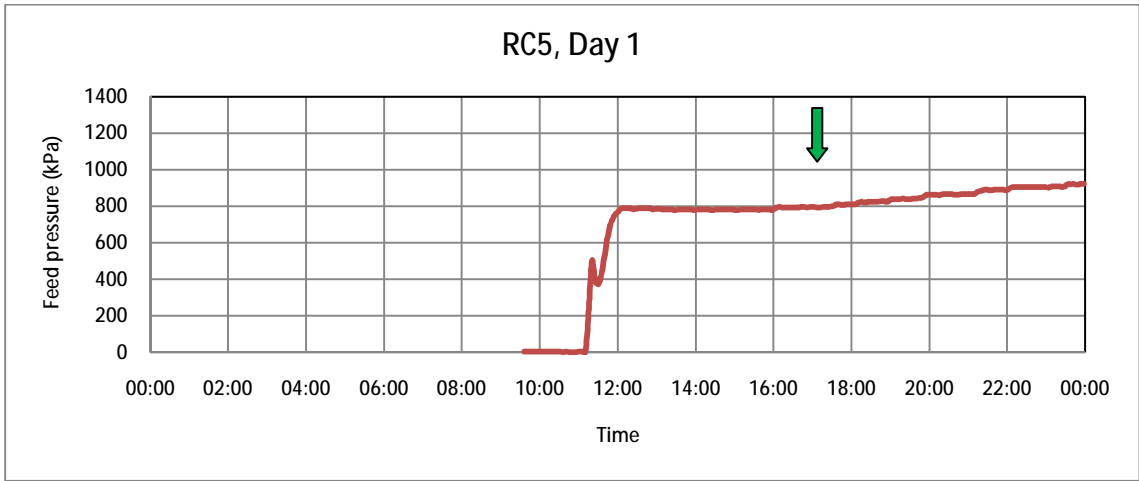


Figure 3.3.2.2.7: RO feed pressure, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

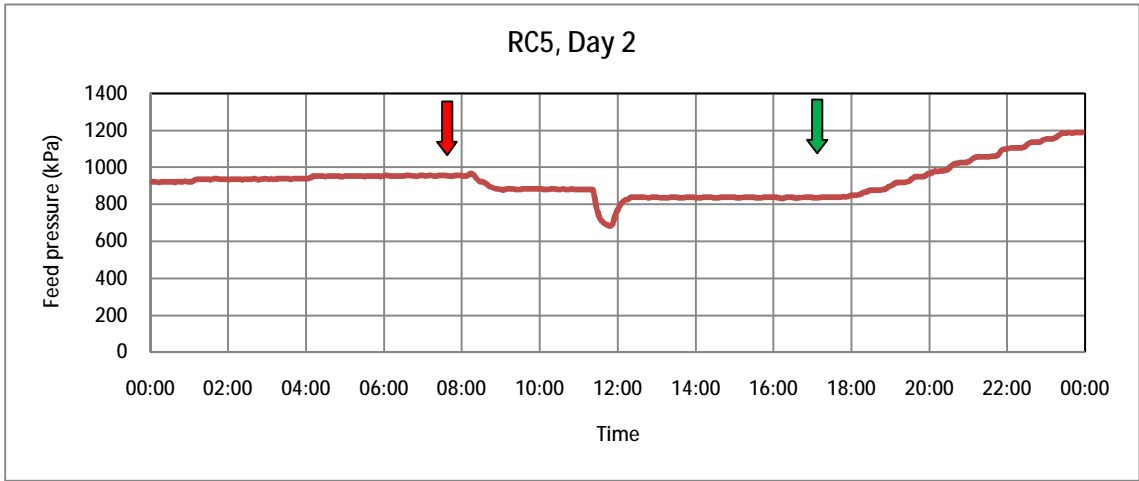


Figure 3.3.2.2.8: RO feed pressure, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

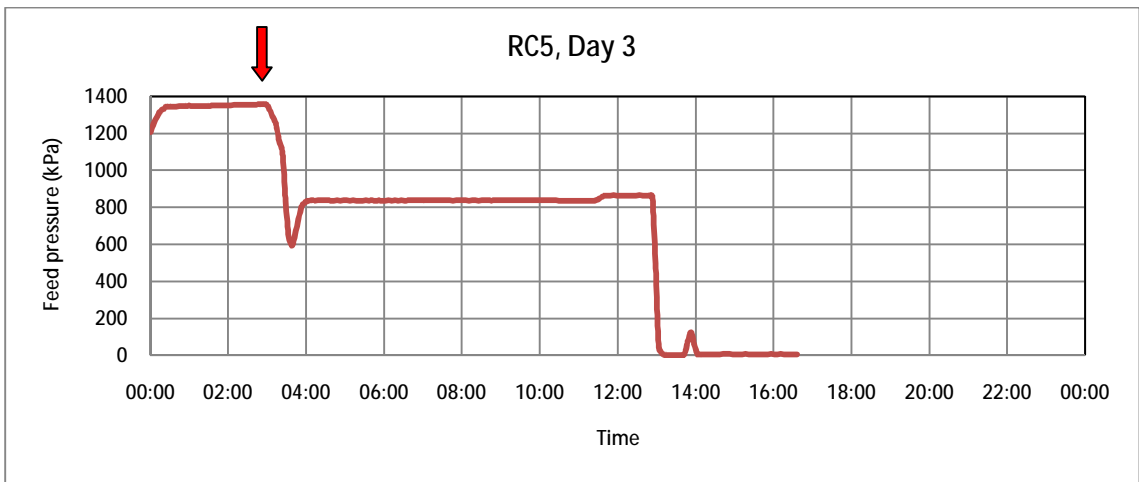


Figure 3.3.2.2.9: RO feed pressure, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

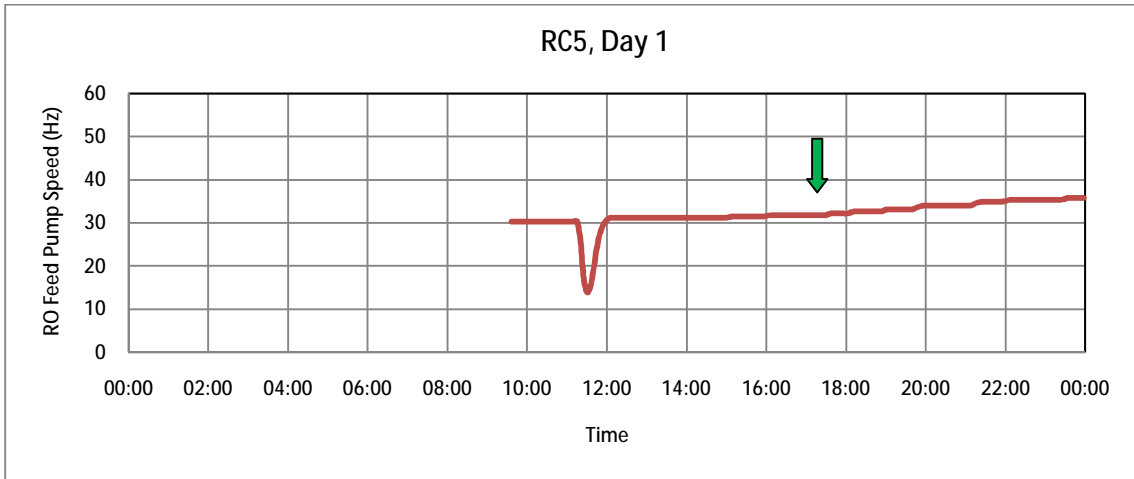


Figure 3.3.2.2.10: RO feed pump speed, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

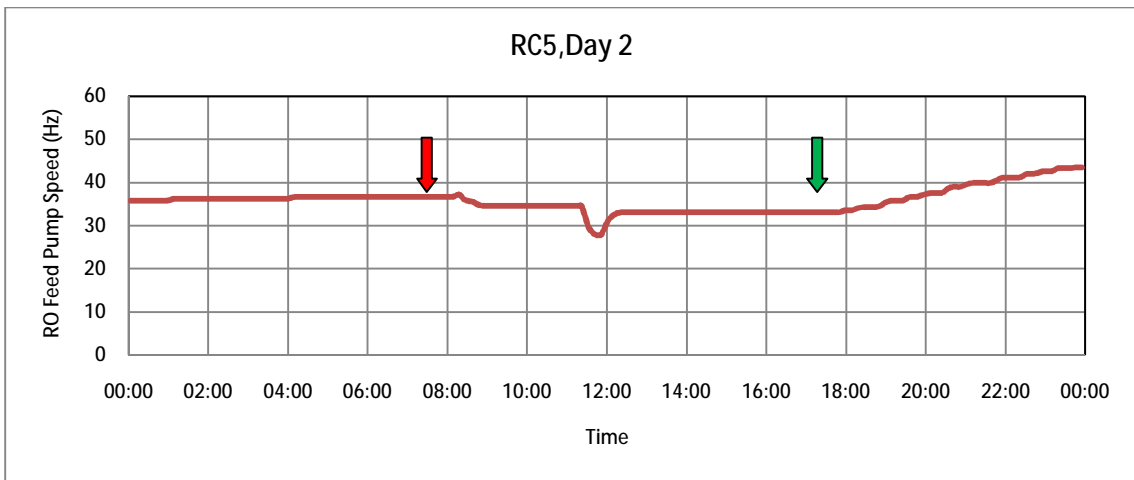


Figure 3.3.2.2.11: RO feed pump speed, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

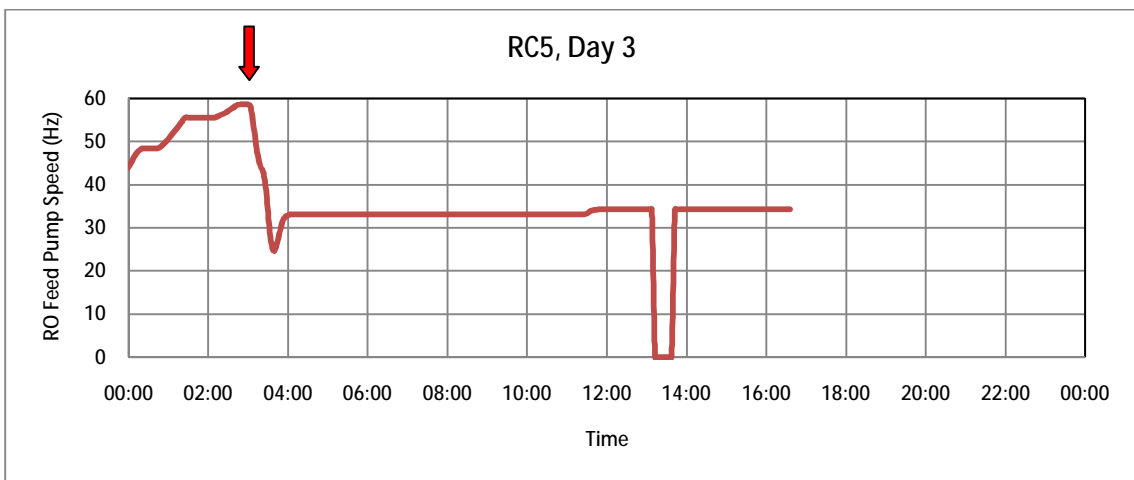


Figure 3.3.2.2.12: RO feed pump speed, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

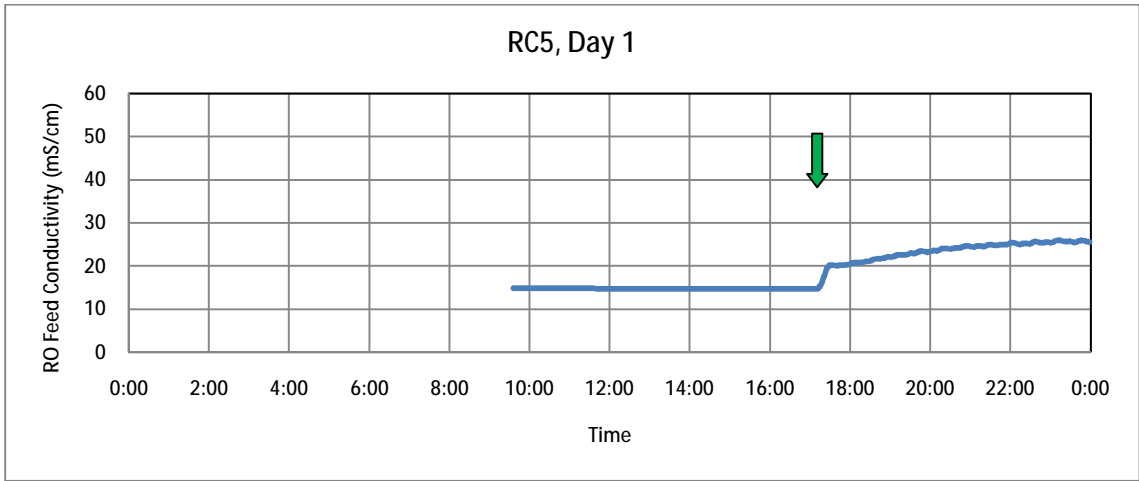


Figure 3.3.2.2.13: RO feed conductivity, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

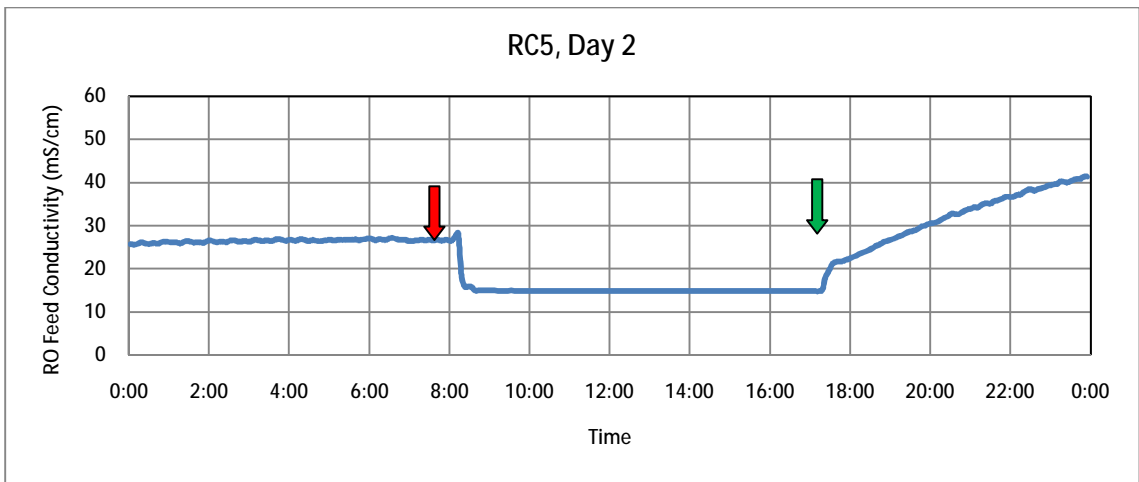


Figure 3.3.2.2.14: RO feed conductivity, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

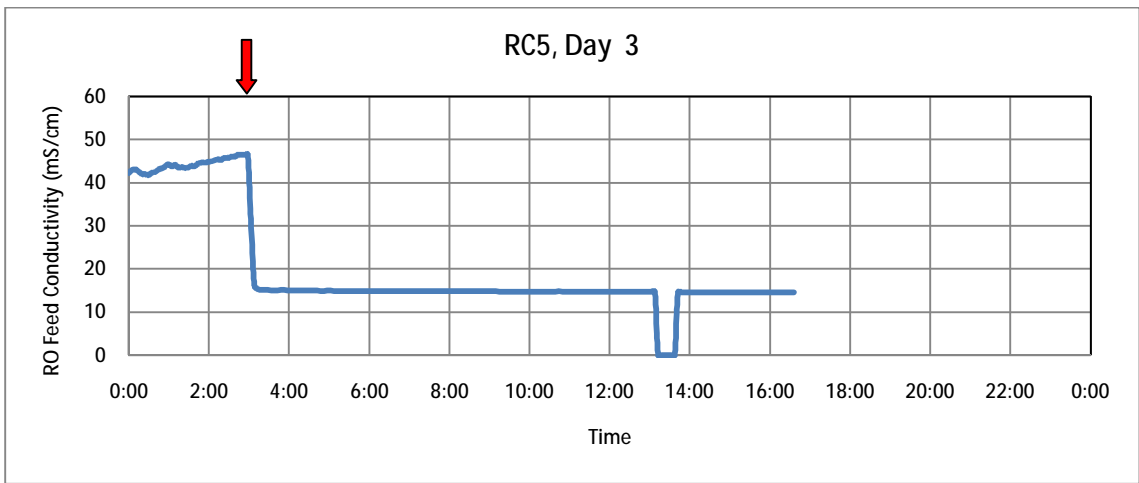


Figure 3.3.2.2.15: RO feed conductivity, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

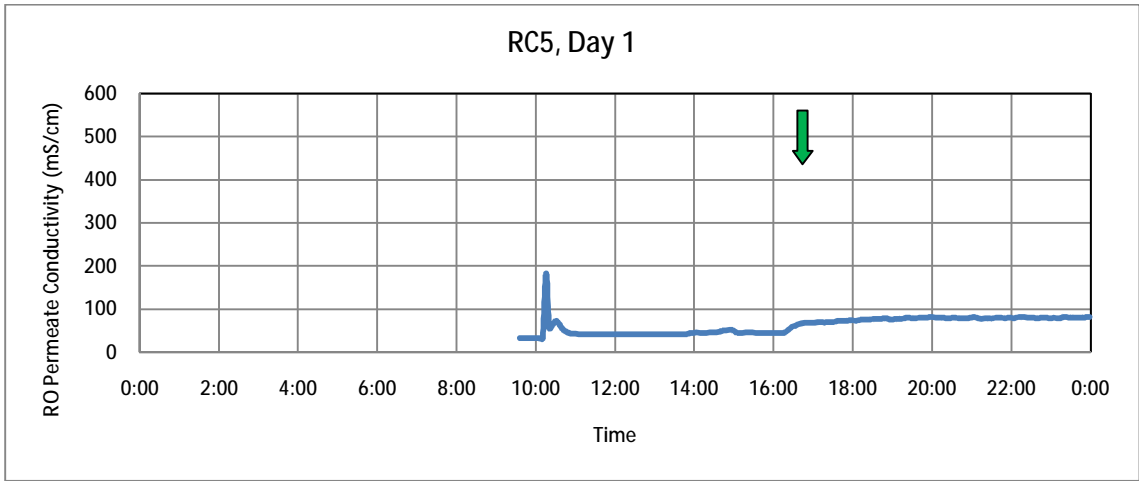


Figure 3.3.2.2.16: RO permeate conductivity, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

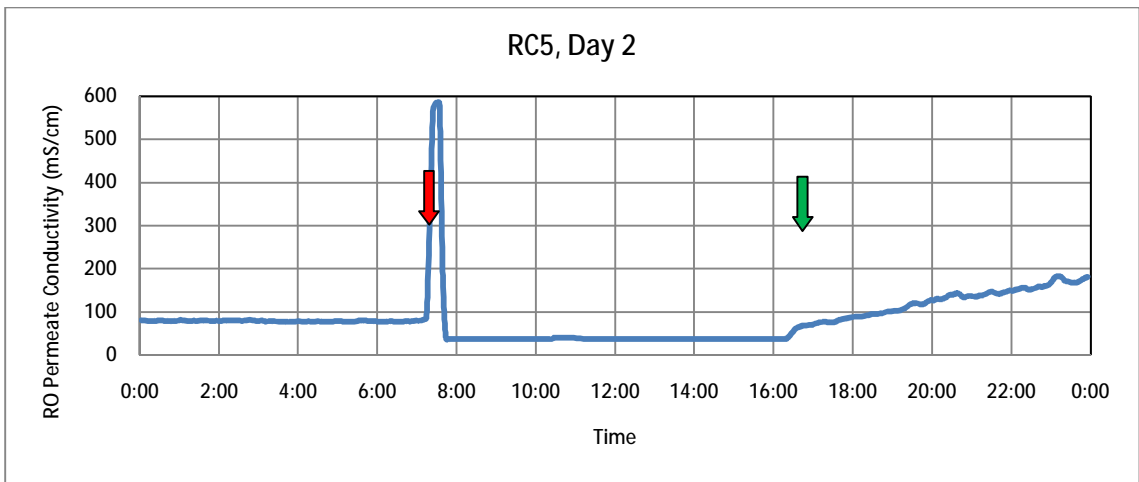


Figure 3.3.2.2.17: RO permeate conductivity, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

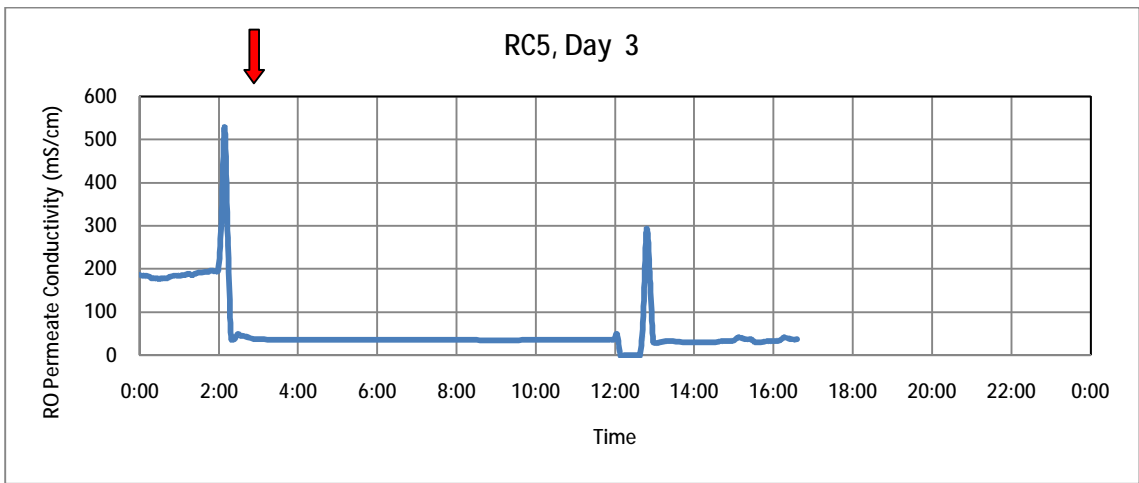


Figure 3.3.2.2.18: RO permeate conductivity, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

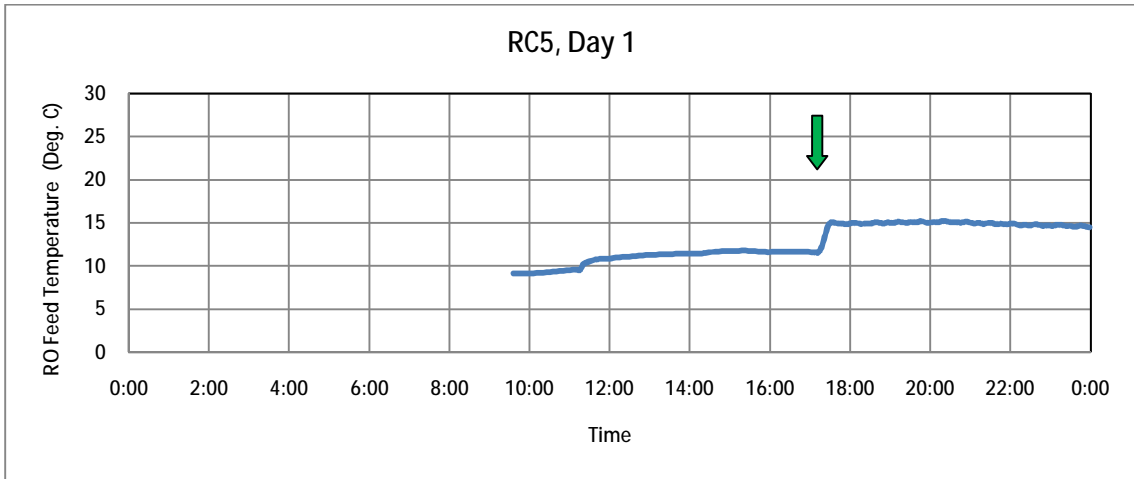


Figure 3.3.2.2.19: RO feed temperature, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by green arrow

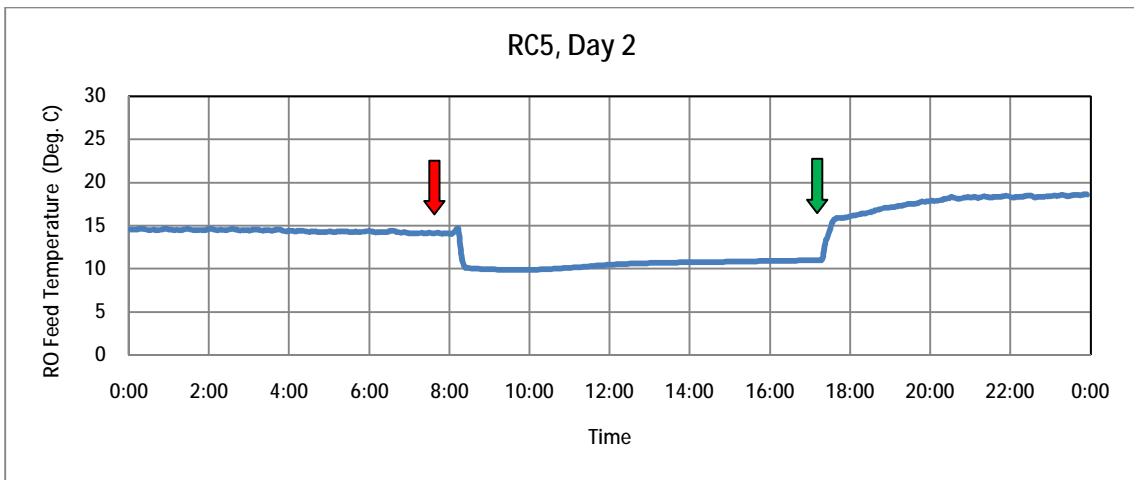


Figure 3.3.2.2.20: RO feed temperature, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

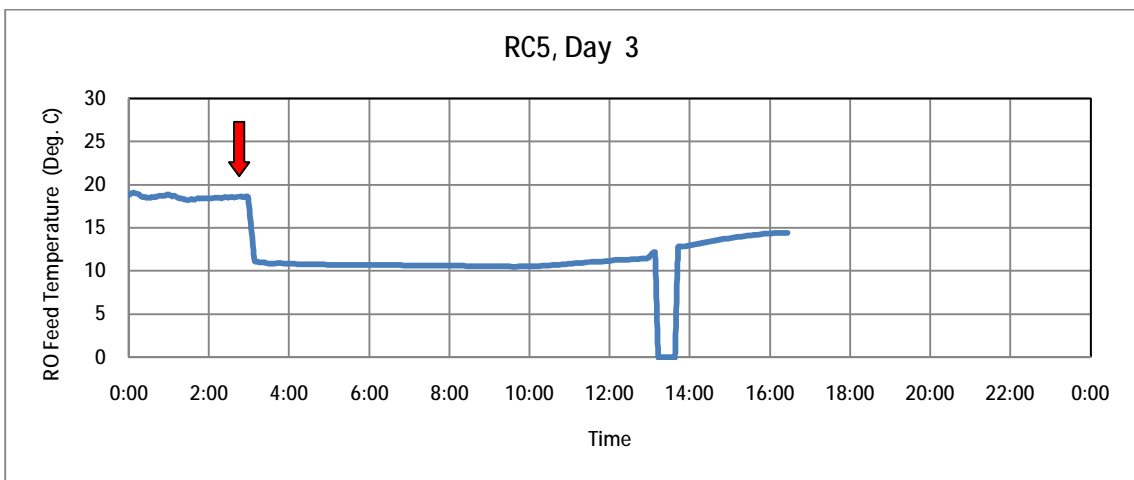


Figure 3.3.2.2.21: RO feed temperature, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

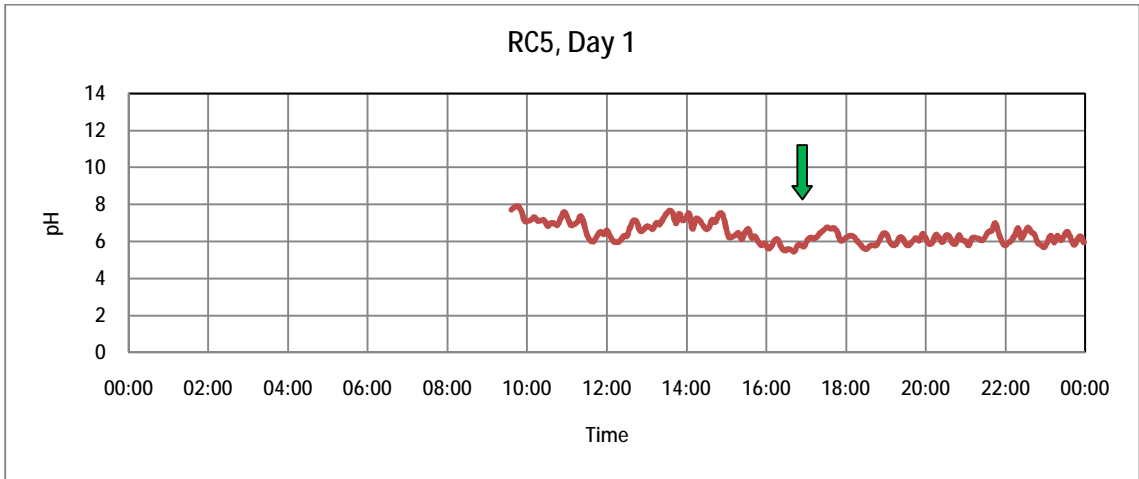


Figure 3.3.2.2.22: RO feed pH, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

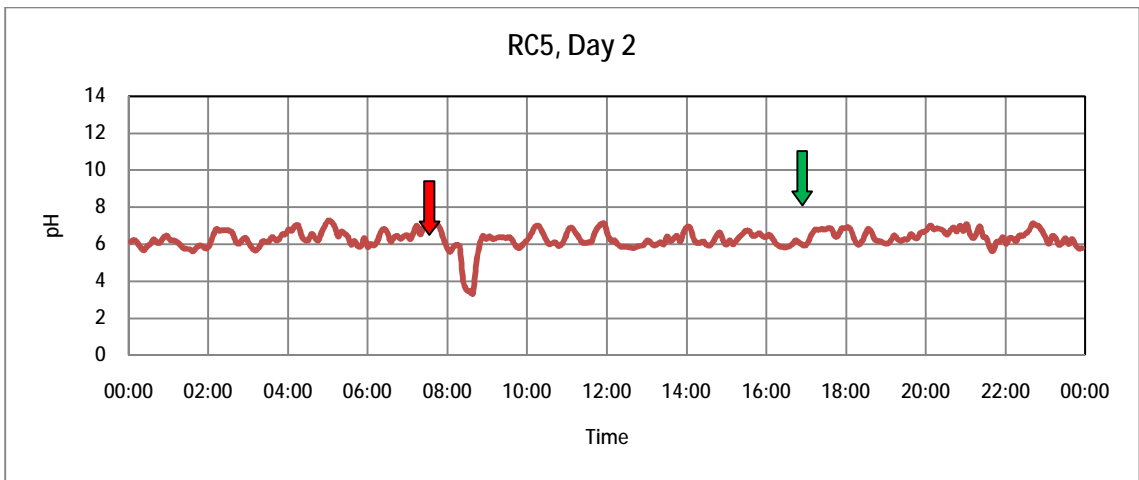


Figure 3.3.2.2.23: RO feed pH, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

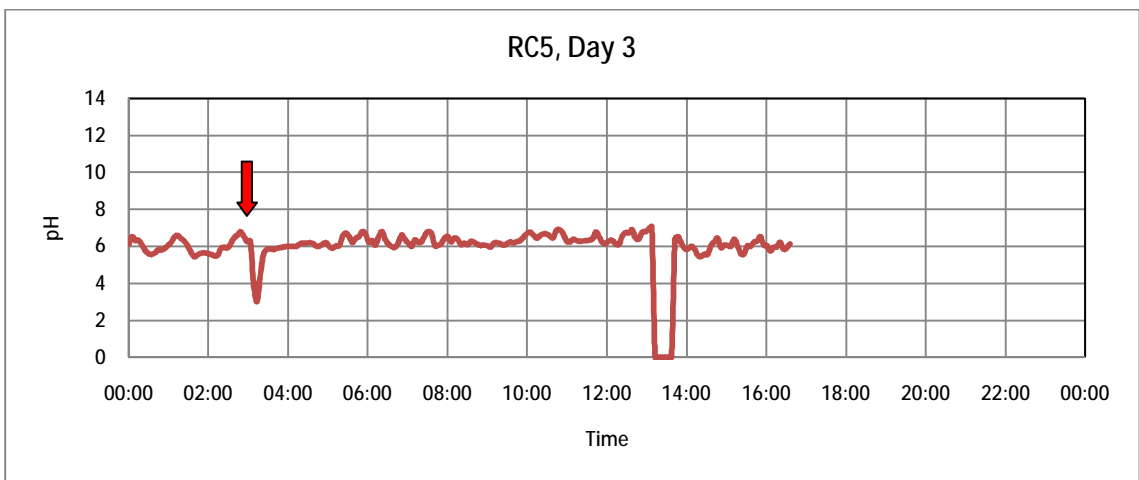


Figure 3.3.2.2.24: RO feed pH, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

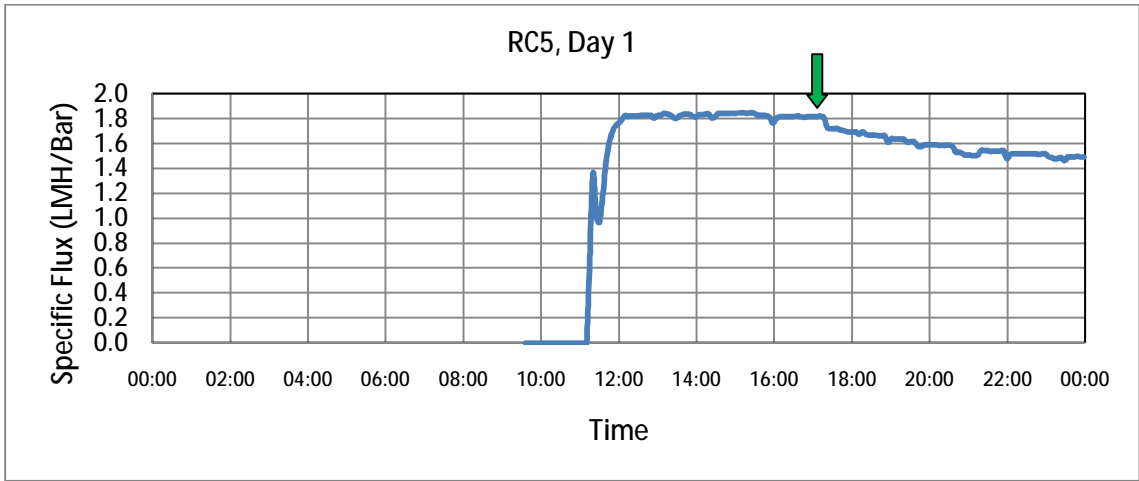


Figure 3.3.2.2.25: Specific permeate flux, RC5, Day 1, connection of the CUF filtrate to the RO feed line is indicated by the green arrow

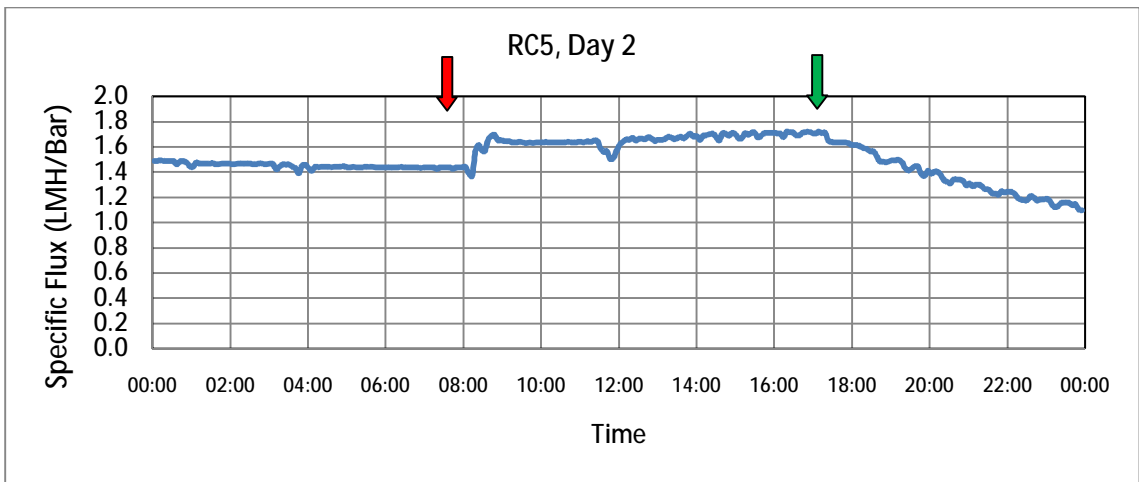


Figure 3.3.2.2.26: Specific permeate flux, RC5, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

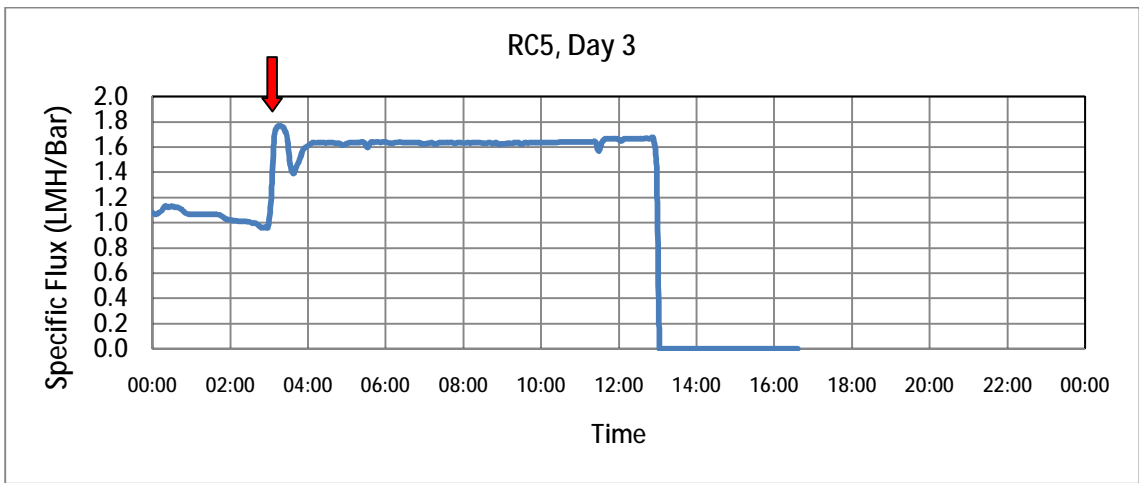


Figure 3.3.2.2.27: Specific permeate flux, RC5, Day 3, disconnection of the CUF filtrate from the RO feed line is indicated by the red arrow

4. Summary and Project Outcomes

4.1 Laboratory scale testing of ASP for Ca reduction

Laboratory trials in the early stages of this project have shown that:

- a) The use of calcium carbonate seed for the accelerated seeded precipitation of calcium from municipal wastewater to achieve high water recoveries is not feasible due to the presence of substances such as magnesium ions and phosphate ions that interfere with calcium precipitation.
- b) The precipitation of calcium as calcium phosphate was found to be less prone to interference
- c) In-situ generation of the seed by addition of phosphate ion at pH 10, followed by reuse of this seed, was an effective means of removal of the calcium.
- d) The quantity of phosphate required in these laboratory experiments was one half to one third of the quantity (in moles) of calcium, at high seed loading rates.

4.2 Pilot scale testing of high recovery RO process

The process was tested on a larger scale in Donald, Victoria, with a pilot plant consisting of dissolved air flotation (DAF), ultrafiltration (UF), reverse osmosis (RO) and accelerated seeded precipitation (ASP) – ceramic ultrafiltration (CUF). A ceramic filtration system was used to remove the seed and precipitated calcium and return the seed to the reaction tanks, and to supply a calcium depleted stream for further water recovery through the RO plant.

4.2.1 Plant design

Pilot Plant reliability

The large number of process steps in the final plant design (DAF-UF-RO-ASP-CUF) made uninterrupted operation of the full plant for more than a few hours very difficult, resulting in a much longer than planned commissioning period.

Pre-treatment Stages

Pre-treatment was initially to be only UF using a pre-existing Hydracap filtration system installed in the United Utilities shipping container. This filtration system was designed to cope with water with turbidity of less than 15 NTU, and it soon became evident in the early stages of the commissioning process that this system was not able to cope with the test water which had a turbidity of 300 to 500 NTU during the trial period due to an algal bloom. This necessitated the installation of a dissolved air flotation plant upstream of the Hydracap UF.

Reverse Osmosis Stage

The major problem encountered with this process stage was the inability of the RO feed pump to achieve the planned water recoveries. Operation at lower feed flow and low RO water recovery was

found to overcome the difficulty associated with the poor pump performance, allowing longer periods of operation, but severely limiting the range of experiments that could be performed.

The brief periods of operation prevent a definitive assessment of whether scaling occurred. The specific flux calculations for the operation of the RO immediately before and after connection to the CUF filtrate stream, however, revealed an 8% decrease in specific flux, suggesting scaling may have taken place. This scaling may be attributable to the formation of calcium phosphate on the membranes due to the excess of phosphate leaving the ASP stage. (see ASP stage and Ceramic Filtration Stage discussions below).

Accelerated Seeded Precipitation Stage

The ASP stage was found to be effective in decreasing the calcium concentration down to less than 8 mg/L from an initial concentration of approximately 380 mg/L. This was, however, at a large excess of phosphate (HPO_4 to Ca mole ratio = 4.0). Investigation of the effect of reduction of this dose was prevented by an intermittent increase in filtrate turbidity, necessitating disconnection of CUF filtrate line from the RO (see Ceramic Filtration Stage discussion below).

Monitoring of calcium levels was performed using a selective calcium ion electrode. Calibration of this electrode using calcium standards with background electrolyte concentrations matching those of the 70% brine used in the laboratory investigations was found to yield results that were within 2 mg/L of the results obtained by atomic adsorption spectroscopy.

Ceramic filtration Stage

Turbidity of Filtrate: The unexpected increase in CUF filtrate turbidity was tentatively attributed to:

- a. The heating in the ceramic filter recirculation loop due to pump energy input
- b. The generation of a supersaturated calcium solution resulting from the decreased solubility of sparingly soluble calcium salts at the higher-temperature [4]. The slow precipitation of sparingly soluble calcium compounds to the higher-temperature solubility limit once the sparingly soluble calcium salt precursor ions reach the filtrate side of the ceramic filter.

This problem may be mitigated by the use of heat exchangers or operation at lower recirculation loop flows that would be expected to generate less heat. This latter mitigation measure may, however, lead to fouling of the ceramic filter due to a higher tendency for cake formation. Another possible mode of mitigation of this problem could be to run the CUF system at higher feed flows, thereby minimising the heating of the solution. This was not an option in the current trials due to the inability of the RO feed pump to achieve high feed flows (see Reverse Osmosis Stage discussion above).

Fouling: Only slight adjustment of the CUF filtrate flow valves was necessary during the trial, indicating that the fouling tendency was low for the ceramic filter at the high recirculation flow and relatively low solids loading (up to ~6 g/L after 24 hours) operating conditions.

4.2.2 Preliminary Cost Estimates

4.2.2.1 Chemical Costs associated with ASP Process Stage

The chemical costs associated with the ASP process stage are shown in Table 4.2.3.1. It can be seen from these calculations that the chemical running costs for the ASP stage are high. The first row of this table is based on the laboratory data presented in Milestone 3 report and represents the best that could be achieved. The calculations in second row of the table are based on the data from the un-optimised running of the pilot plant, and these figures represent the expected worst case scenario.

Table 4.2.2.1 ASP stage chemical cost per 1000 L of product water

	Cost of required phosphate per 1000 L of feedwater *	Cost of required caustic per 1000L of feedwater #	Total Cost per 1000L of feedwater	Total Cost per 1000L of product water
	(\$)	(\$)	(\$)	(\$)
Laboratory Trial Data	1.53	0.07	1.60	1.78
Pilot Plant trial Data (not optimised)	18.7	0.07	18.8	20.9

* Ca content of feedwater is 220 mg/L (5.5 mole/1000L), Laboratory trial mole phosphate requirement = calcium mole content of feedwater / 3, Pilot plant trial mole phosphate requirement = calcium mole content of feedwater x 4, cost of phosphate = \$0.84/mole

caustic requirement to raise pH to 10 = 50 mL of 46% Caustic/1000L of RO concentrate, = 24 mL of 46% Caustic/1000L of feedwater , cost of caustic solution = \$2.83/L of 46% caustic solution.

The above costs may be mitigated by the value of generated calcium phosphate. Calcium phosphate is commonly used as fertiliser and the cost of crude grade calcium phosphate is approximately \$0.8/kg.

To produce 1 kL of product water at 90% water recovery, 1.111 kL of wastewater must be treated. The quantity of calcium precipitated from 1111 L of 200 mg/L Ca wastewater = $200 \times 1111 \text{ mg} = 222 \text{ g of Ca}$. This equates to $755 \text{ g of CaHPO}_4 (= 222 \times ((40 + 96)/40))$.

The desalination process could generate approximately \$ 0.6 worth of calcium phosphate from every kL of product water produced.

4.2.2.2 Energy requirements

The energy requirements of the high recovery RO process are considerably greater than those of associated with conventional RO (see Table 4.2.2.2). Not only is there the added energy associated with the running of the ceramic filtration system, but also the added energy associated with the extraction of water from a more concentrated brine. The high recovery RO energy requirement is also very large as it applies to the worst case scenario, where the RO is run at low water recovery and approximately equal quantities of raw water is blended with the calcium depleted process stream.

It is important to note, however, that RC5 on which the energy calculations were performed was inefficient due to the high recycling rate through the RO (see Figure 3.3.2.1). Running the process at higher RO water recoveries results in less recycling through the RO and greater energy efficiencies for the RO stage. The Filmtec Modelling software, ROSA, predicts that the specific energy for RC4, for example, would be 3.9 kWh/ kL of product water.

It is also important to note that the specific energy for the CUF system is expected to be less at a larger scale (e.g. 4.5 to 5 kWh/kL at the ML scale).

Table 4.2.2.2: Energy costs associated with desalination stage, per kL of product water (pre-treatment not included)

	Conventional RO (75% water recovery)	High Recovery RO (90% water recovery)
RO specific energy (kWh/m ³)*	1.5	5.5
CUF specific energy (kWh/m ³)	-	13.5
Total Energy Requirement	1.5	19
Total cost (\$/kL)**	0.3	3.8

* Using ROSA Dow-Filmtech Modelling Software

** At 20c/kwh

4.2.2.3 Comparison of Evaporation pond requirements

The pondage requirement of the 90% RO operation is expected to be one third that of conventional RO operated at 75%. The costs for a desalination plant producing 1 ML/d of product water are shown in Table 4.2.2.3.

Table 4.2.2.3: Estimated evaporation pond capital expenditure costs for a desalination plant producing 1 ML/d product water.

	Conventional RO (75% water recovery)	High Recovery RO (90% water recovery)
Feed flow required to produce 1 ML/d of product water (ML)	1.333	1.111
Volume of waste brine produced (ML/d)	0.333	0.111
Area of evaporation pond required (ha)*	12.15	4.05
Total Cost (\$M)	12.15	4.05

* Assuming an average evaporation rate of 10 ML/ha/yr

** Assuming a cost of \$1M/ha

5. Conclusions and Recommendations

Laboratory trials in the early stages of this project have shown that the removal of calcium ions from the test water by accelerated seeded precipitation (ASP) using calcium carbonate seed was not effective. This was attributed to the presence of substances such as magnesium ions and phosphate ions which interfere with calcium carbonate precipitation. The precipitation of calcium as calcium phosphate, however, was found to be less prone to interference. In-situ generation of seed by addition of phosphate ion at pH 10, followed by reuse of this seed, was found to be an effective means of removal of the calcium. The quantity of phosphate required in these laboratory experiments was one half to one third of the quantity (in moles) of calcium.

The pilot plant used in these studies consisted of dissolved air flotation, ultrafiltration, reverse osmosis, accelerated seeded precipitation and ceramic ultrafiltration. The large number of process steps in the final plant design made uninterrupted operation of the full plant for more than a few hours very difficult, resulting in a much longer than planned commissioning period. Despite this major difficulty, however, the plant was operated at 90% water recovery for long enough for the major impediment to its operation and to the prolonged achievement of high water recovery to manifest itself – turbidity forming after the ceramic ultrafiltration (CUF) system that prevented further RO processing to extract more clean water.

Running of the pilot plant was characterised by unexpected increases in CUF filtrate turbidity above the 1 NTU requirement for the RO process. The generation of a turbid CUF filtrate was tentatively attributed to the heating in the ceramic filter recirculation loop due to pump energy input, giving rise to the generation of a supersaturated calcium solution due to the decreased solubility of sparingly soluble calcium salts at the higher-temperature, and the precipitation of calcium from this supersaturated solution after it passed through the ceramic filter. The increases in CUF turbidity during this trial necessitated the disconnection of the CUF filtrate from the RO, thereby reducing run time and preventing investigations which require longer time periods, such as seed dose and chemical optimisation, and RO scale formation assessments.

The brief periods of operation prevented a definitive assessment of whether fouling of the RO membranes due to scale formation occurred. The specific flux calculations for the operation of the RO immediately before and after connection to the CUF filtrate stream, however, revealed an 8% decrease in specific flux, suggesting scaling may have taken place. This scaling may be due to formation of calcium phosphate on the membranes due to the excess of phosphate used during the trial. This may be mitigated by optimisation of phosphate dosing. This was not performed due to short run time resulting from the generation of turbid CUF filtrate.

Only slight adjustment of the CUF filtrate flow valves was necessary during the trial, indicating that the fouling tendency was low for the ceramic filter at the high recirculation flow and relatively low solids loading (up to ~6 g/L after 24 hours) operating conditions used.

The use of a calcium selective electrode was found to a convenient and reliable method of monitoring the calcium concentration.

Preliminary cost estimates for the ASP process indicate that the ASP-RO process considerably more expensive than conventional RO, but that there are considerable cost savings associated with lower

evaporative pond required to manage the waste brine. The chemical running costs associated with the ASP stage suggested from the laboratory experiments were \$1.8 per kL of product water. The un-optimised pilot plant ASP stage chemical usage costs were \$21 per kL of product water. These chemical running costs may be mitigated by the use of the generated calcium phosphate as fertiliser. The value of the generated calcium phosphate was calculated to be \$0.7/kL of product water. The energy running costs of the plant were estimated to be \$3.8 /kL of product water, which is 13 times the energy cost of conventional RO run at 75% water recovery. At an equal product water output, the cost of evaporation pond construction at 90% water recovery were estimated to be one third those of conventional RO run at 75% water recovery.

Recommendations for Further Research

The removal of the seed and precipitated calcium is central to the success of the process, and further research is required to determine the best method of removal of these solids. The chemistry of calcium phosphate precipitation appeared to vary throughout these trials because of changes in temperature, and thus the design of the solid-liquid separation process in the ASP was unable to achieve the required low turbidity values under all conditions. The kinetics of calcium phosphate precipitation as a function of temperature and the mode of generation of turbidity needs to be investigated. It is only after an effective and reliable mode of separation of the seed and precipitated calcium salts from the treated RO concentrate is found that the seed material can be allowed to build up to useful concentrations (~20 g/L according to laboratory tests) and optimisation of process inputs such as chemicals and energy can be made. Other modes of solids separation, such as sedimentation, should also be considered for this application.

The detection of the incidence of scale formation relies on comparisons of flux and pressure that are best performed at constant feedwater composition, and further research in this area would be facilitated by the use of two smaller RO plants rather than one larger one. The recirculation of the calcium-depleted ASP-treated RO concentrate back to the same RO that had generated the concentrate in this trial resulted in long periods during which the feed water to the RO was changing in composition, making early detection of scale formation very difficult.

Future research projects in this area should focus exclusively on the ASP, solids removal and downstream RO processing, without having to in any way be concerned with pre-treatment and the generation of the RO concentrate that is to be feedwater for the ASP process. A willing partner that already performs the pre-treatment and generates a concentrate should be sought and the pilot ASP research facilities should be set up at the source of the RO concentrate that is to be feedwater for the ASP process. The cost of chemicals and energy for this research can be minimised by constructing and operating a smaller ASP plant.

Although optimisation of the chemical and energy inputs was not performed during this trial, it is obvious that the process is likely to be an expensive one for the highly saline (~10 g/L salt) and calcium rich (~0.2 g/L) test water used in this trial. Future research should consider application to less challenging wastewaters.

Potential benefits to local water Industry

Addition of another process stage to the conventional RO desalination process to achieve higher water recoveries is expected to incur an added cost associated with the additional chemicals and energy inputs. The benefits of this added cost, however, are that the quantity of waste brine produced is less and the quantity of product water is more at higher water recoveries. At equal clean water production rates, the evaporation pond requirement for a plant operated at 90% water recovery is one third that of one operated at 75% water recovery. If 95% water recovery could be achieved, the evaporation pond requirement would be reduced to one sixth that at 75% water recovery. This would, however, require considerably more energy than operation at 90% water recovery, and much more than operating at 75% water recovery.

Economic justification for aiming to achieve higher water recoveries can only be made if the benefit of the lower evaporation pond requirement outweighs the higher chemical and energy running costs. This is most likely to be the case in locations where there is a desperate need for fresh water and where land costs and regulatory requirements make the construction of large evaporation ponds prohibitively expensive.

Acknowledgements

The authors wish to thank Dow Filmtec for donation of the SW membrane elements used in this research

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- (2) Anditya Rahardianto, Junbo Gao, Christopher J. Gabelich, Mark D. Williams, Yoram Cohena, High recovery membrane desalting of low-salinity brackish water: Integration of accelerated precipitation softening with membrane RO, *Journal of Membrane Science* 289 (2007) 123–137.
- (3) Anditya Rahardianto, Brian C. McCool, Yoram Cohen, Accelerated desupersaturation of reverse osmosis concentrate by chemically-enhanced seeded precipitation, *Desalination* 264 (2010) 256–267.
- (4) Yonghui Song, Hermann H. Hahn, Erhard Hoffmann, Effects of solution conditions on the precipitation of phosphate for recovery: A thermodynamic evaluation, *Chemosphere* 48 (2002) 1029–1034.

Appendix:

Commissioning Results: RC1 to RC4

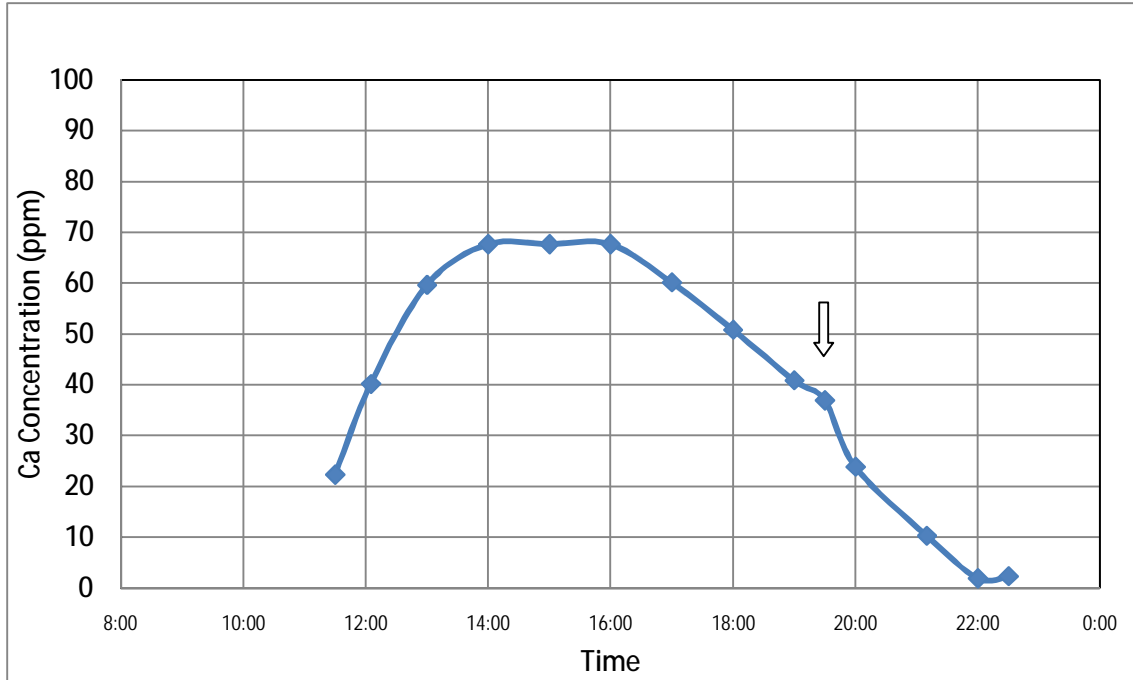


Figure 3.2.1.1: Calcium concentration in ceramic filter filtrate, run condition 1 (RC1). connection of the CUF filtrate to the RO feed line is indicated by the arrow

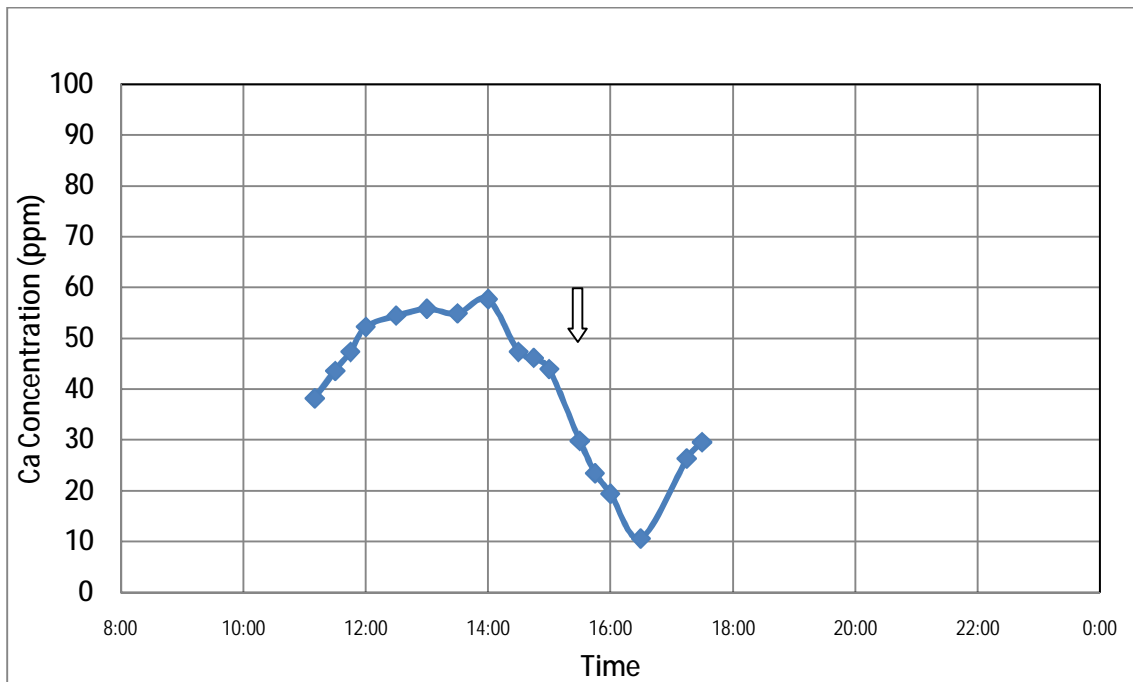


Figure 3.2.1.2: Calcium concentration in ceramic filter filtrate, run condition 2 (RC2). connection of the CUF filtrate to the RO feed line is indicated by the arrow

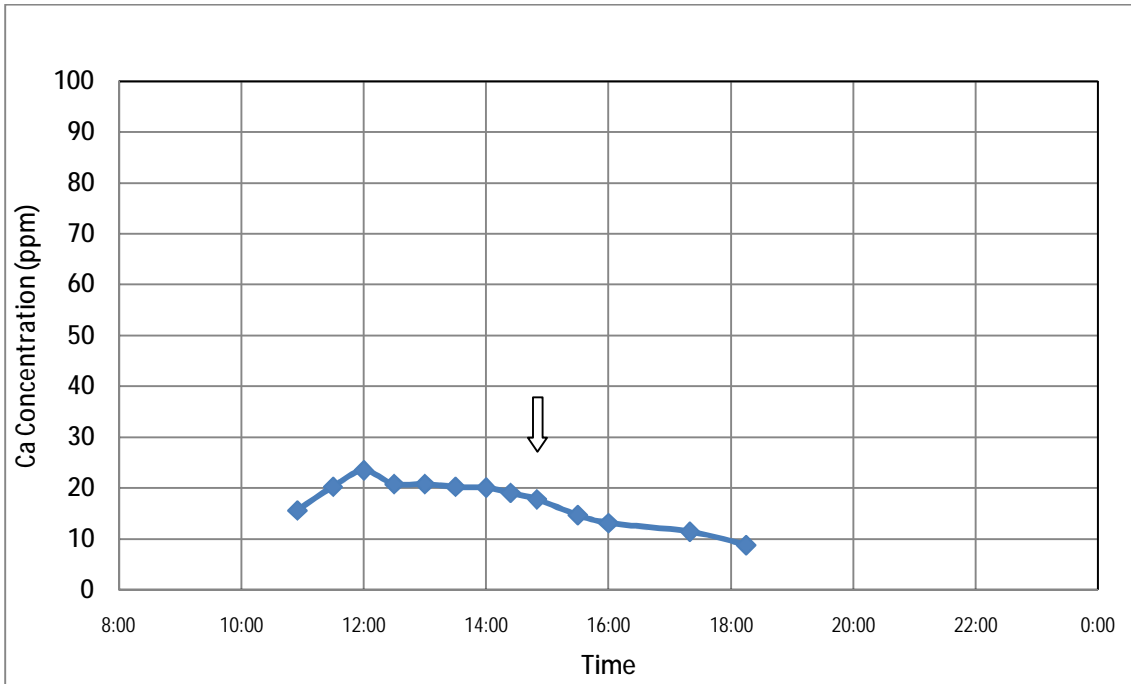


Figure 3.2.1.3: Calcium concentration in ceramic filter filtrate, run condition 3 (RC3). connection of the CUF filtrate to the RO feed line is indicated by the arrow

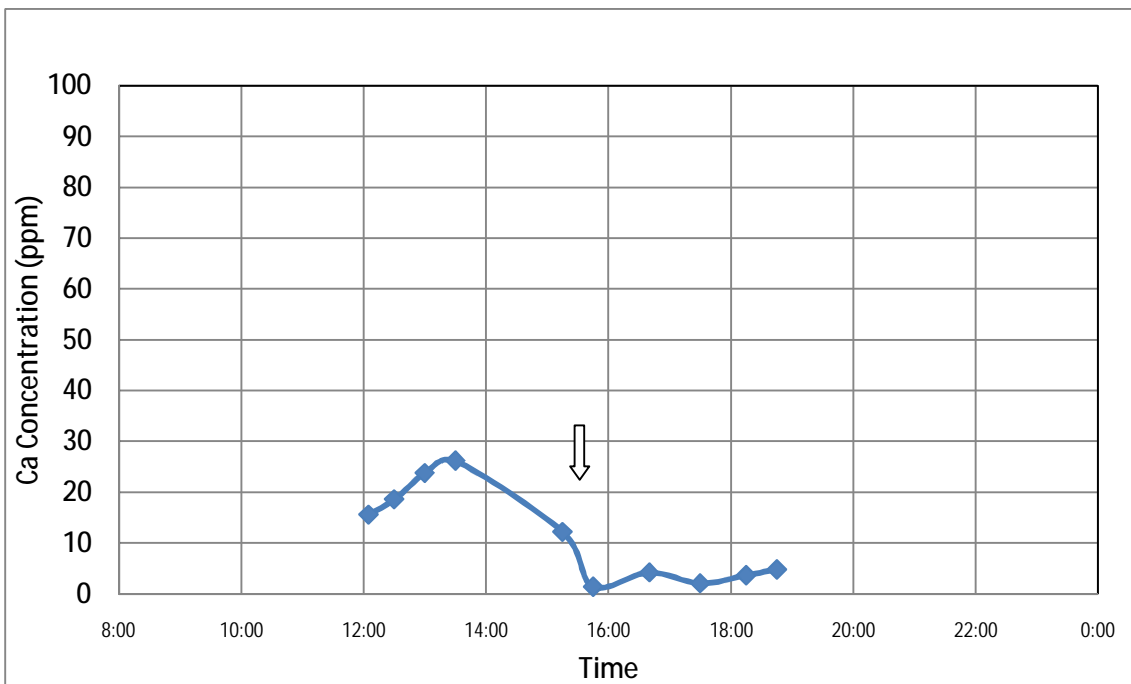
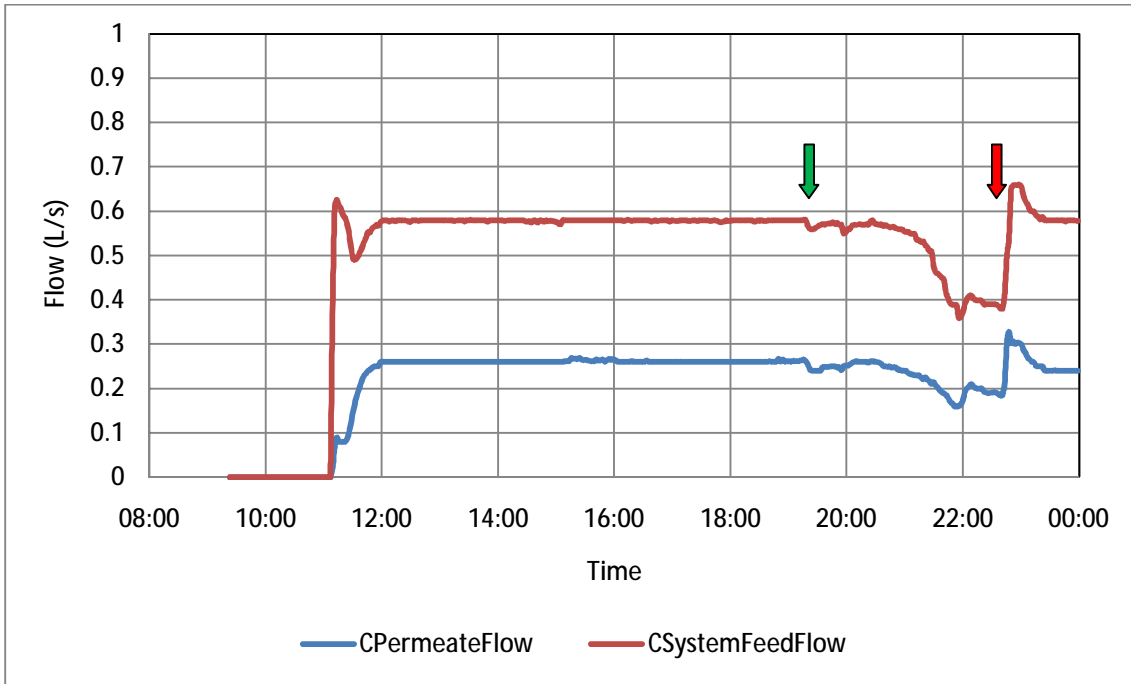
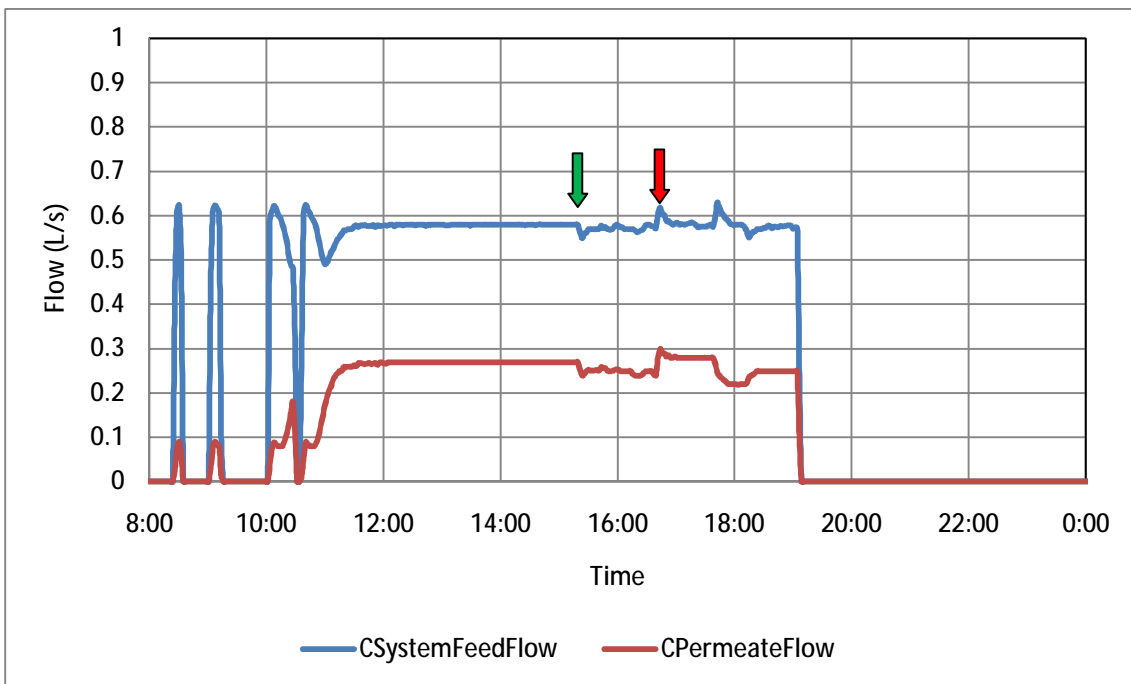


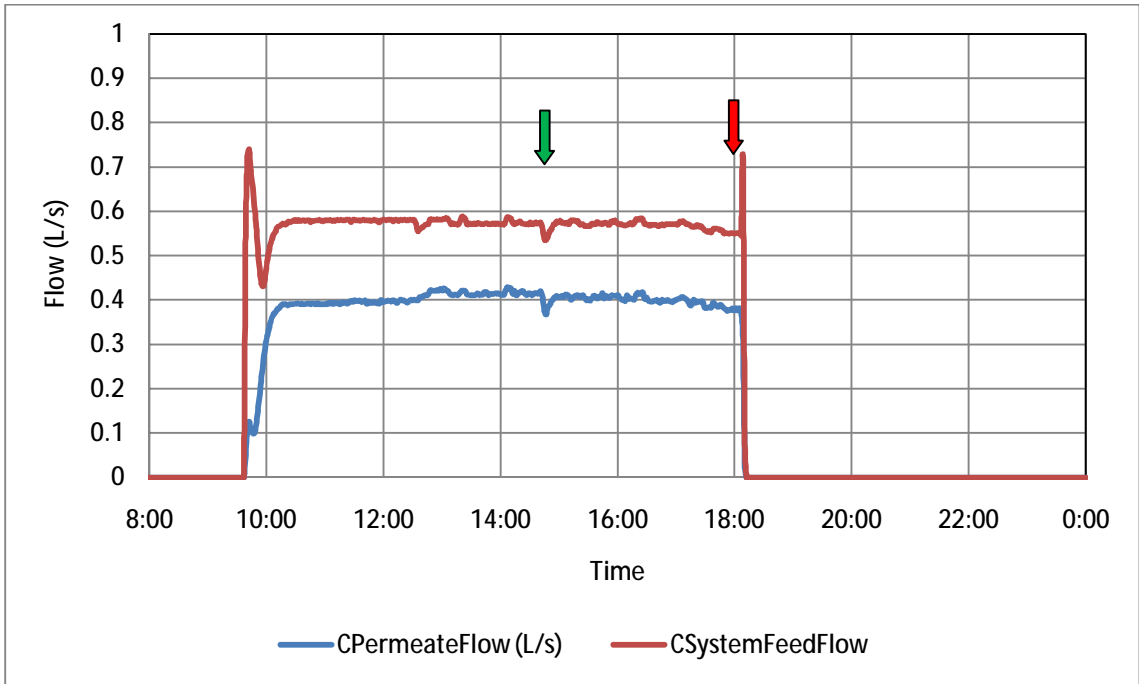
Figure 3.2.1.4: Calcium concentration in ceramic filter filtrate, run condition 4 (RC4). connection of the CUF filtrate to the RO feed line is indicated by the arrow



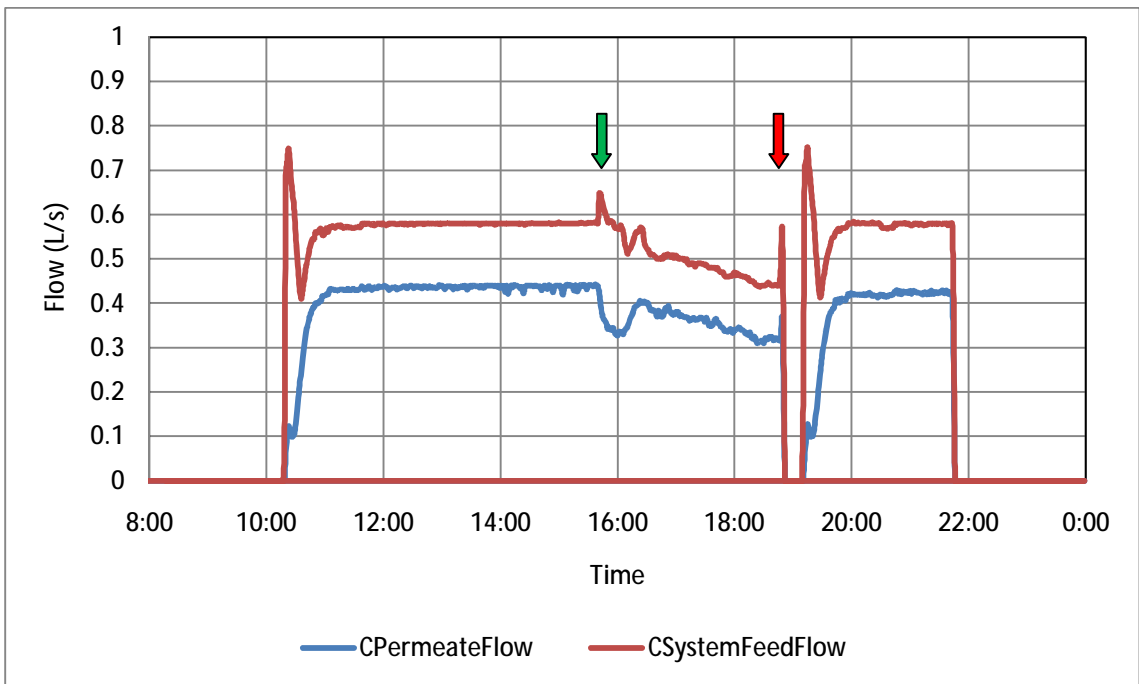
Figures 3.2.2.1: Feed flow and permeate flow, RC1 preliminary run, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively



Figures 3.2.2.2: Feed flow and permeate flow, RC2 preliminary run, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively



Figures 3.2.2.3: Feed flow and permeate flow, RC3 preliminary run, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively



Figures 3.2.2.4: Feed flow and permeate flow, RC4 preliminary run, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

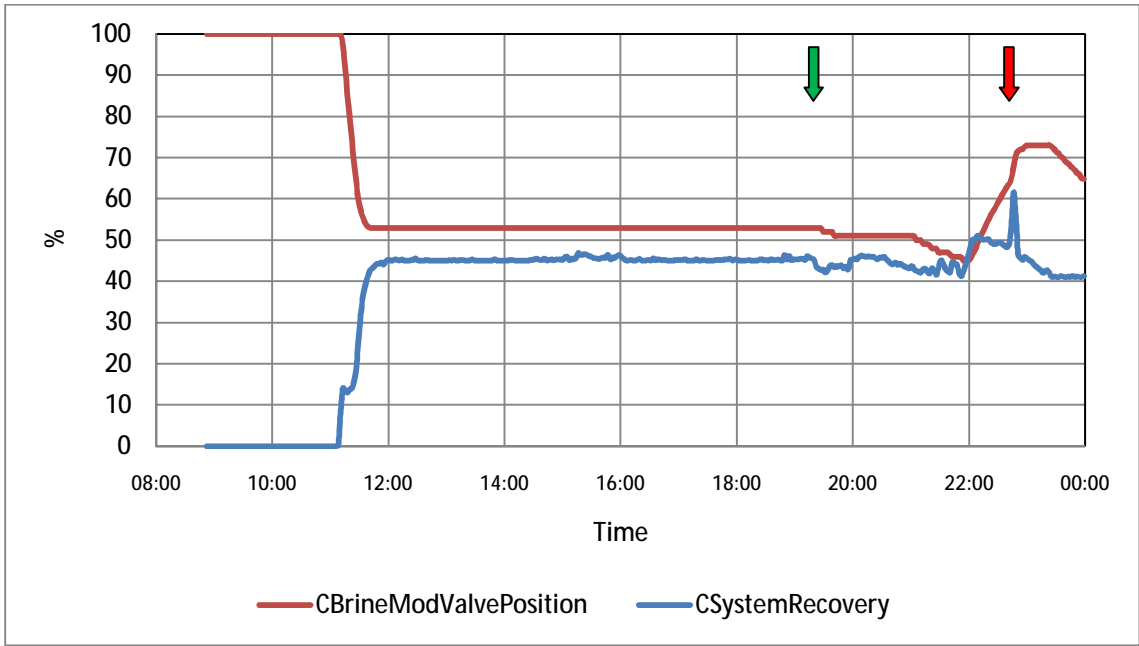


Figure 3.2.2.5: RO water recovery and brine constriction valve position, RC1, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

(compare to Figure 3.2.2.1)

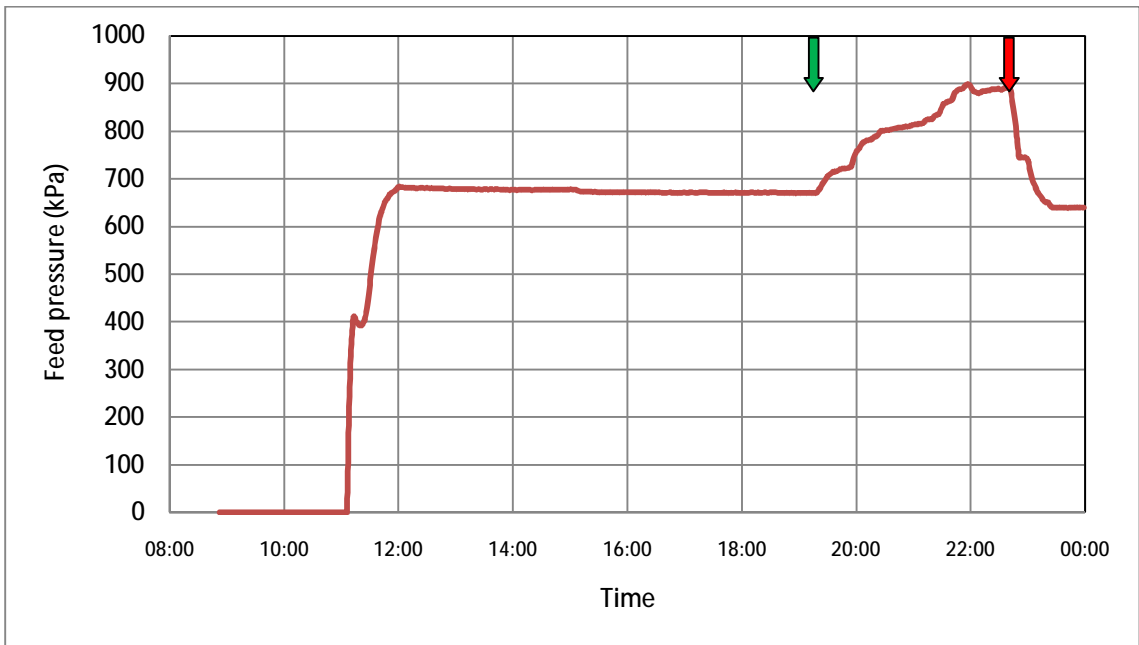


Figure 3.2.2.6: RO feed pressure, RC1, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

(compare to Figure 3.2.2.1).

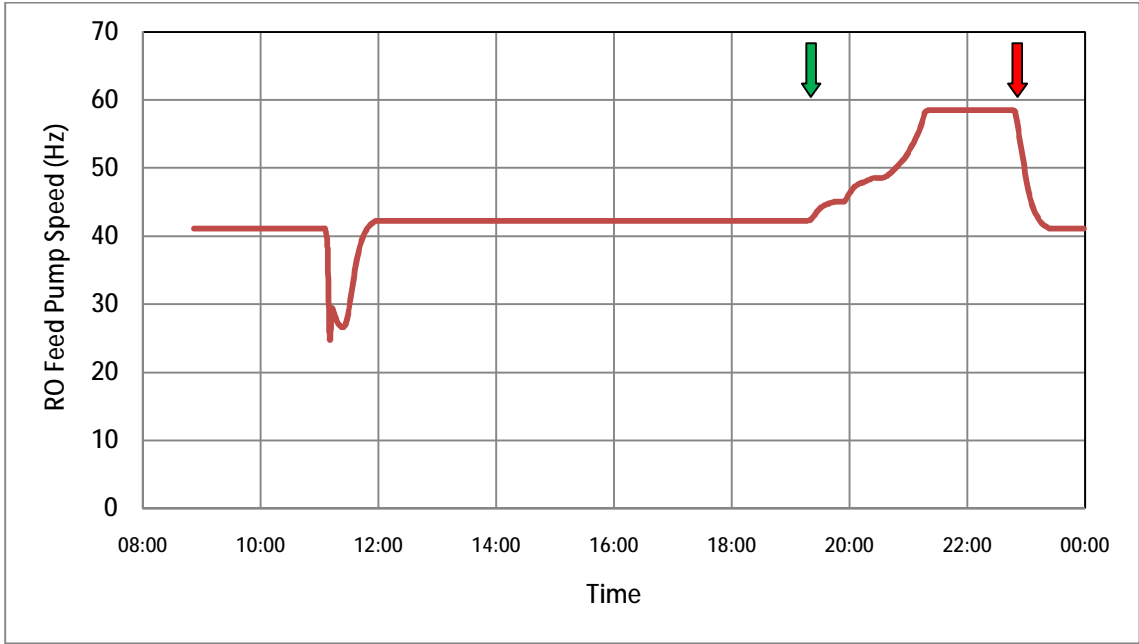


Figure 3.2.2.7: RO feed pump speed, RC1, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

(compare to Figure 3.2.2.1)

RC2 Experiment results

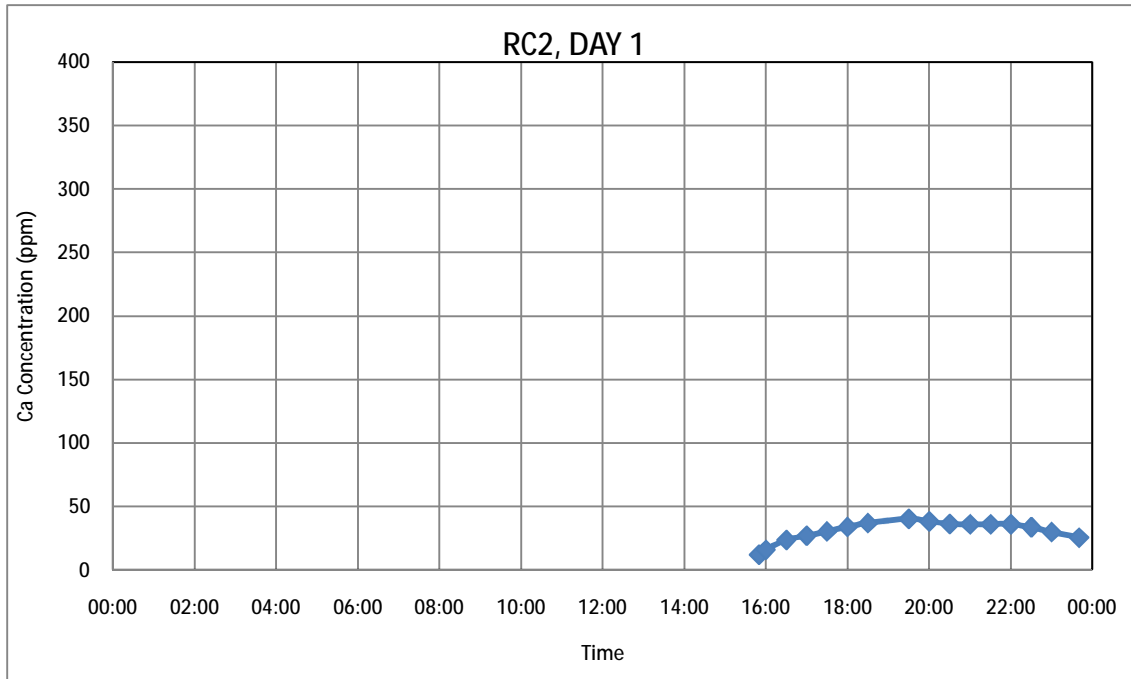


Figure 3.3.1.1.1: CUF filtrate calcium concentration, RC2, Day 1

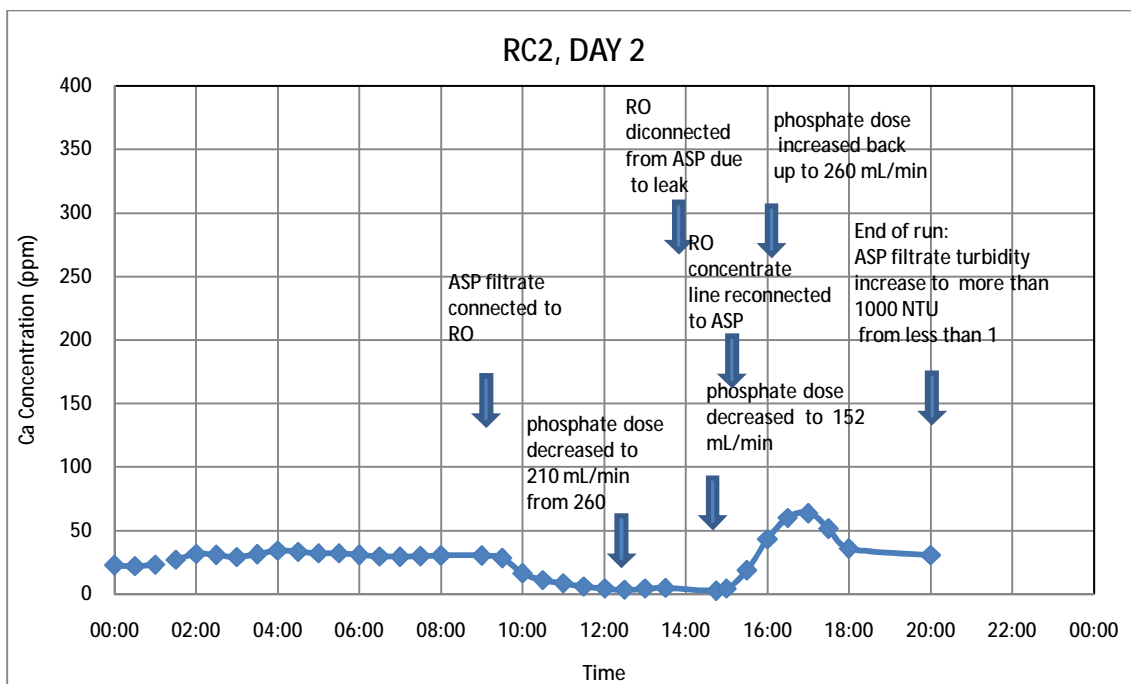


Figure 3.3.1.1.2: CUF filtrate calcium concentration, RC2, Day 2

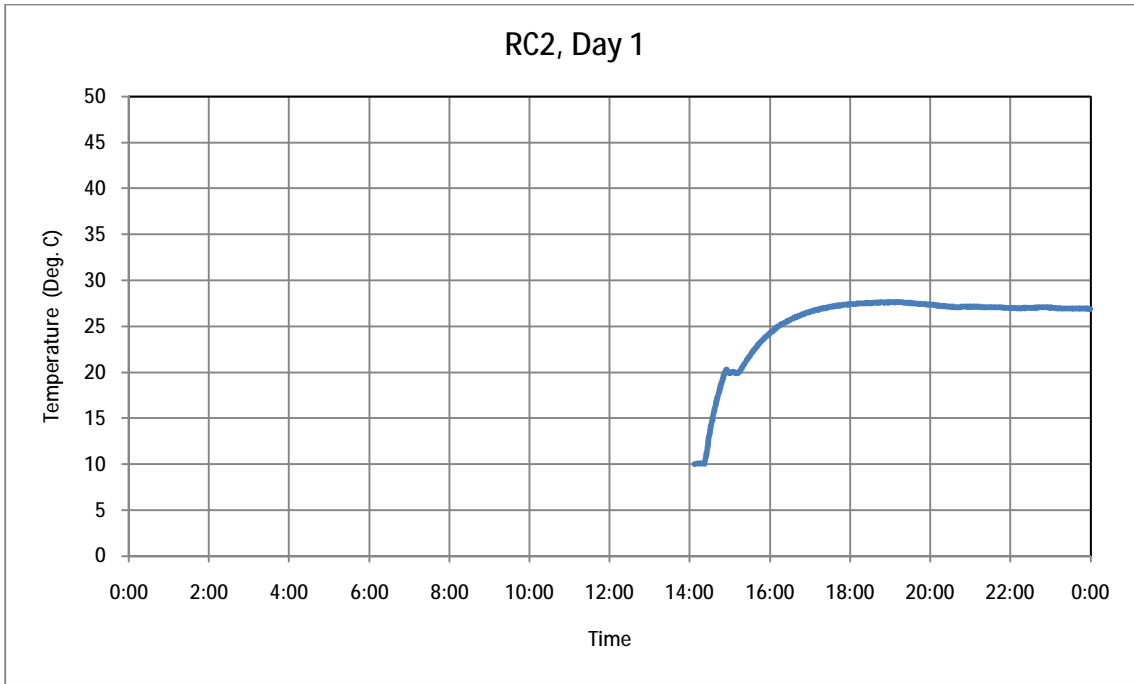


Figure 3.3.1.1.3: CUF filtrate temperature, RC2, Day 1

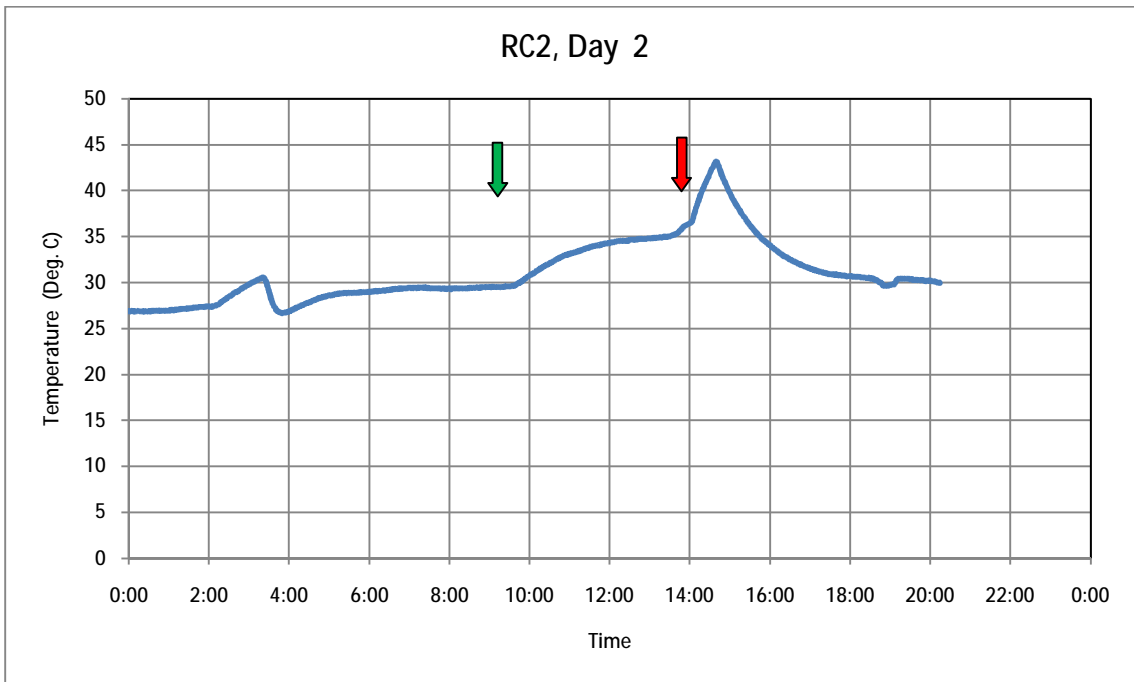


Figure 3.3.1.1.4: CUF filtrate temperature, RC2, Day 2

Table 3.3.1.1.1: Ceramic filter filtrate turbidity results, RC2

Run Day	Time	Turbidity
1	18:50	0.27
	19:00	0.22
	20:30	0.15
	21:00	0.22
	21:30	0.15
	22:00	0.14
	22:30	0.15
	23:00	0.21
	23:40	0.77
2	0:01	0.17
	0:30	0.17
	1:00	0.17
	1:30	0.42
	2:00	0.2
	2:30	0.18
	3:00	0.21
	3:30	0.2
	4:00	0.17
	4:30	0.17
	5:00	0.17
	5:30	0.17
	6:00	0.15
	6:30	0.19
	7:00	0.17
	7:30	0.14
	8:00	0.16
	14:50	0.68
20:00	0.2	
21:10	404	
21:15	1000	

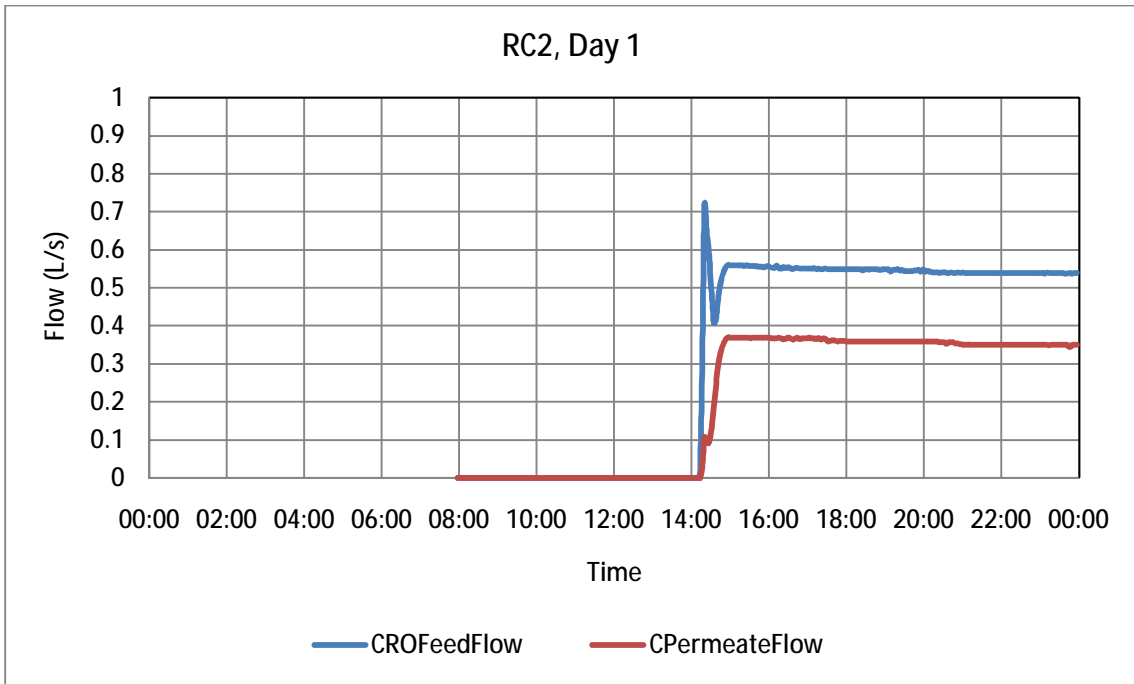


Figure 3.3.1.2.1: RO feed flow and permeate flow, RC2, Day 1

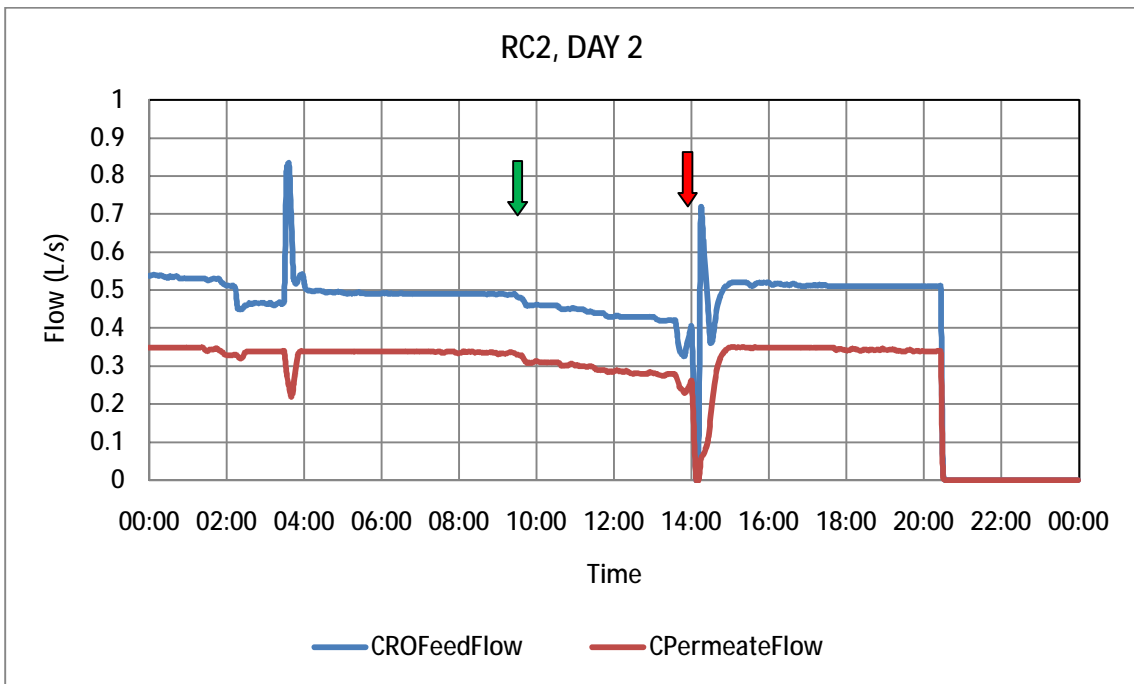


Figure 3.3.1.2.2: RO feed flow and permeate flow, RC2, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

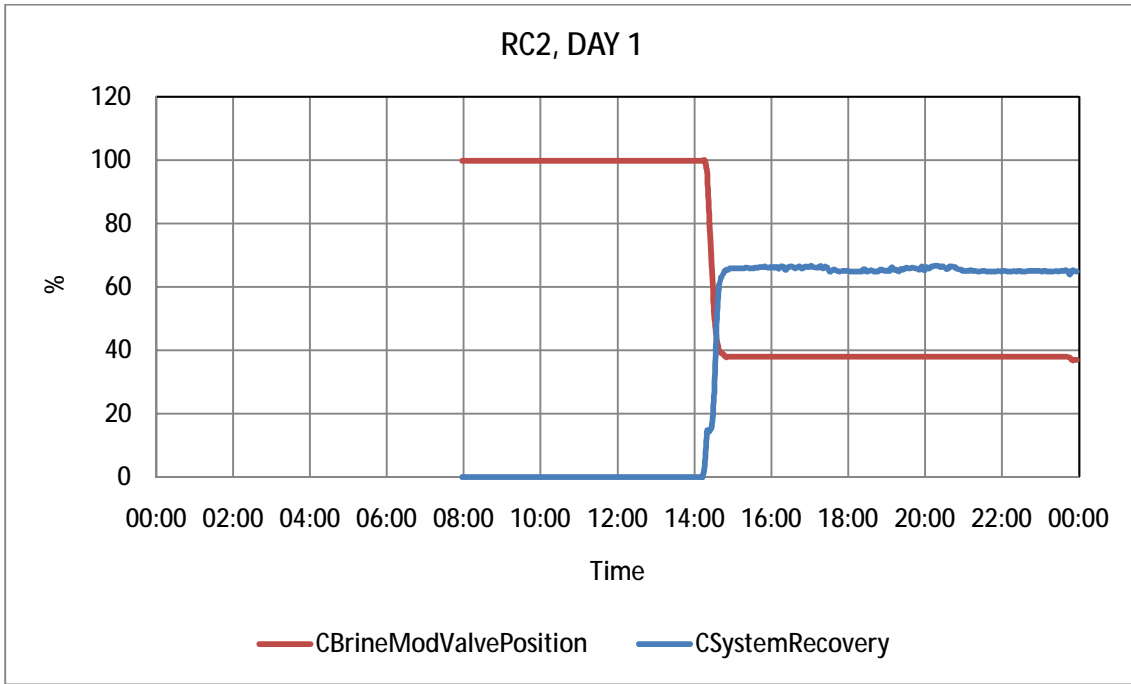


Figure 3.3.1.2.3: RO water recovery and brine constriction valve position, RC2, Day 1

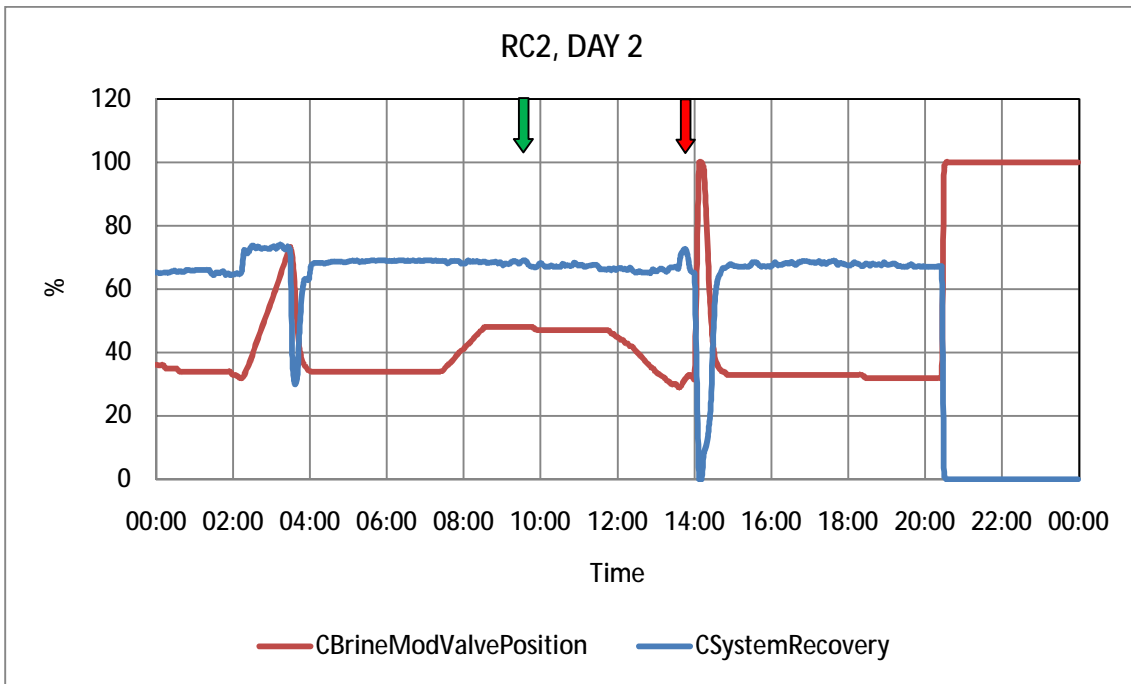


Figure 3.3.1.2.4: RO water recovery and brine constriction valve position, RC2, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

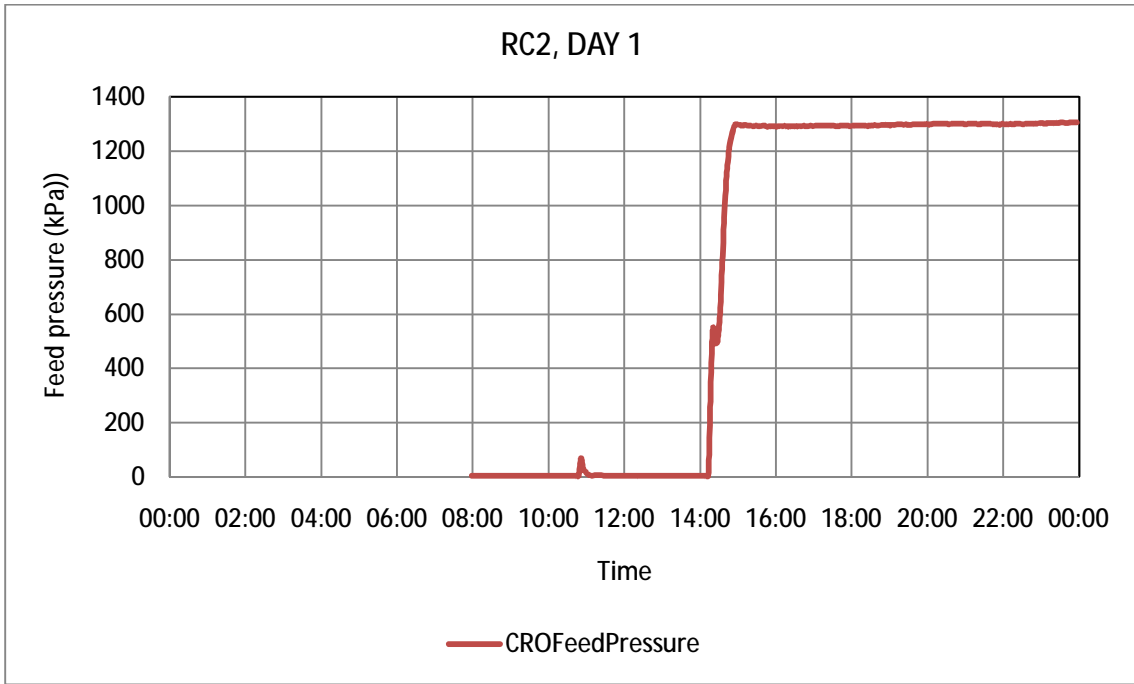


Figure 3.3.1.2.5: RO feed pressure, RC2, Day 1

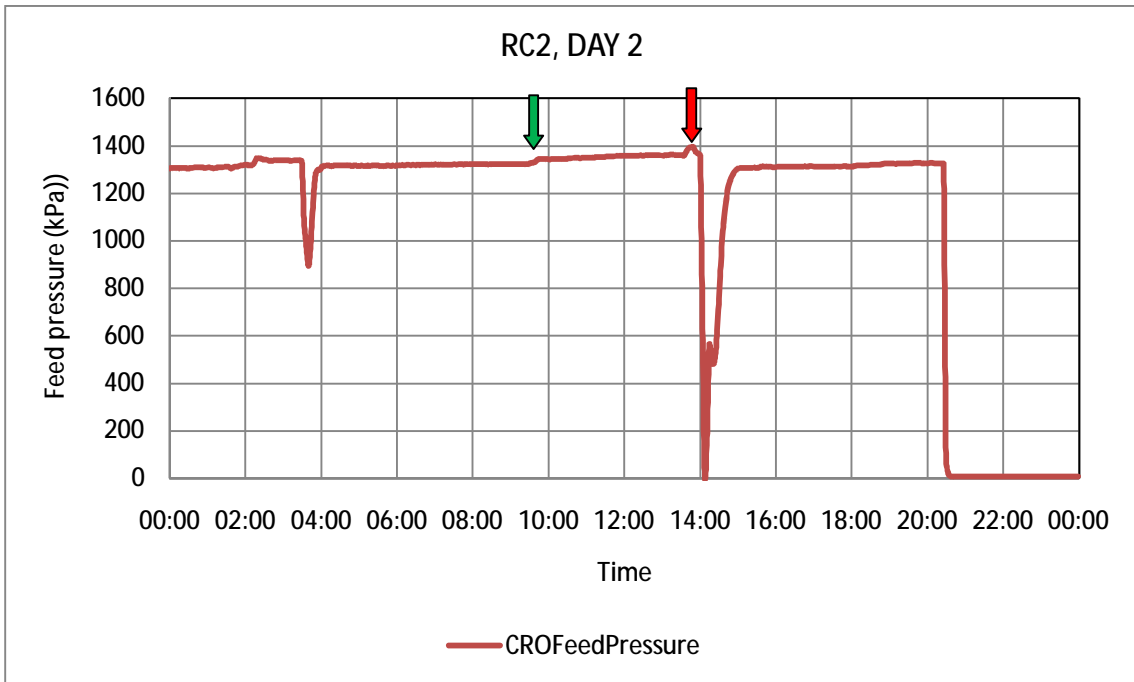


Figure 3.3.1.2.6: RO feed pressure, RC2, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

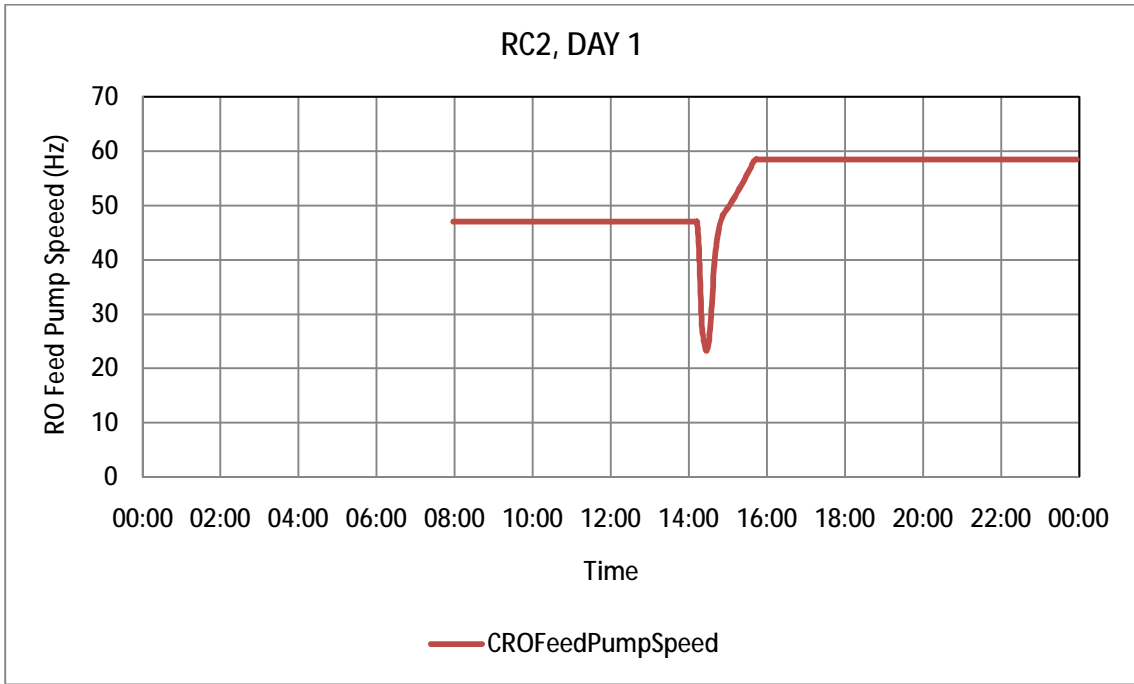


Figure 3.3.1.2.7: RO feed pump speed, RC2, Day 1

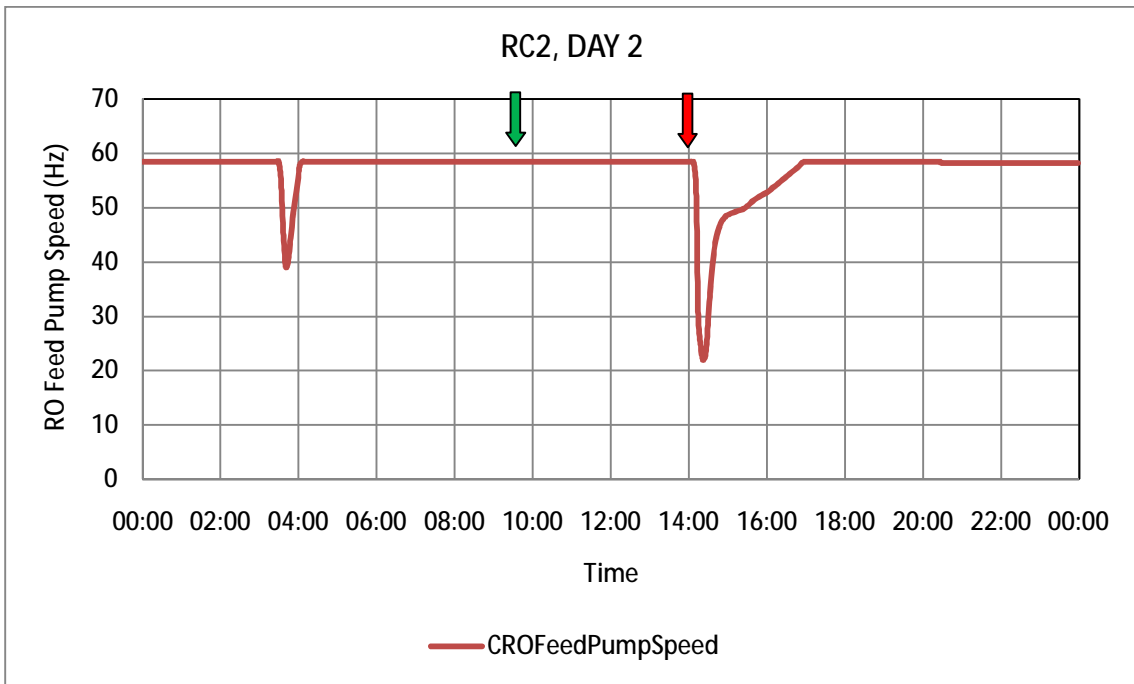


Figure 3.3.1.2.8: RO feed pump speed, RC2, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

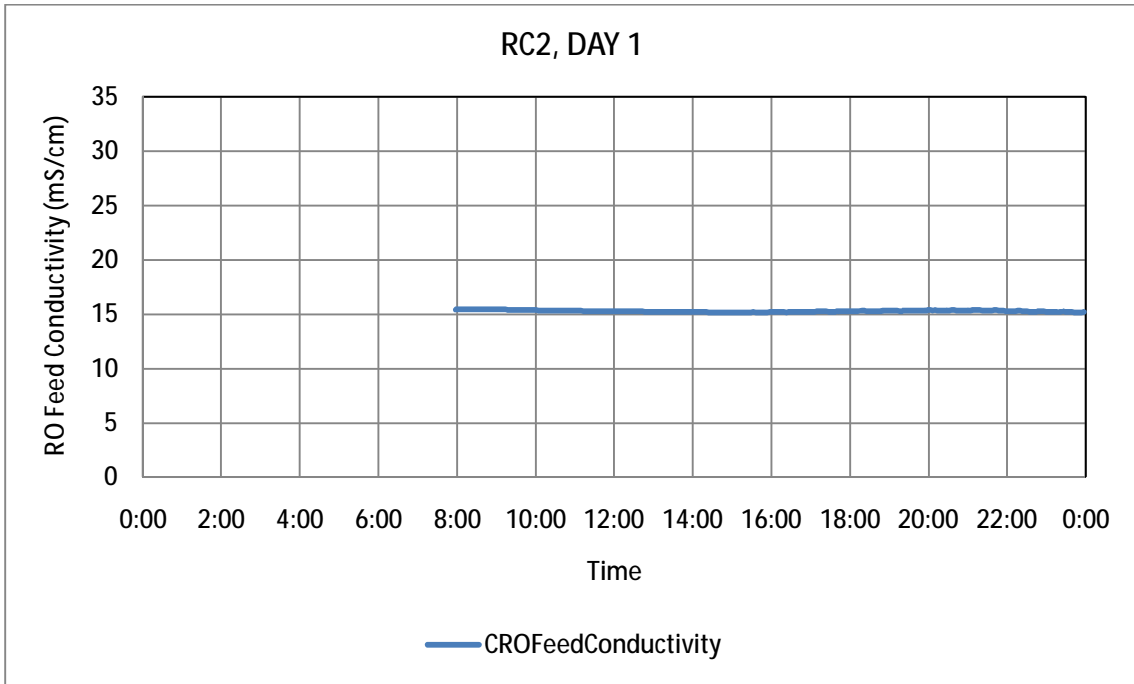


Figure 3.3.1.2.9: RO feed conductivity, RC2, Day 1

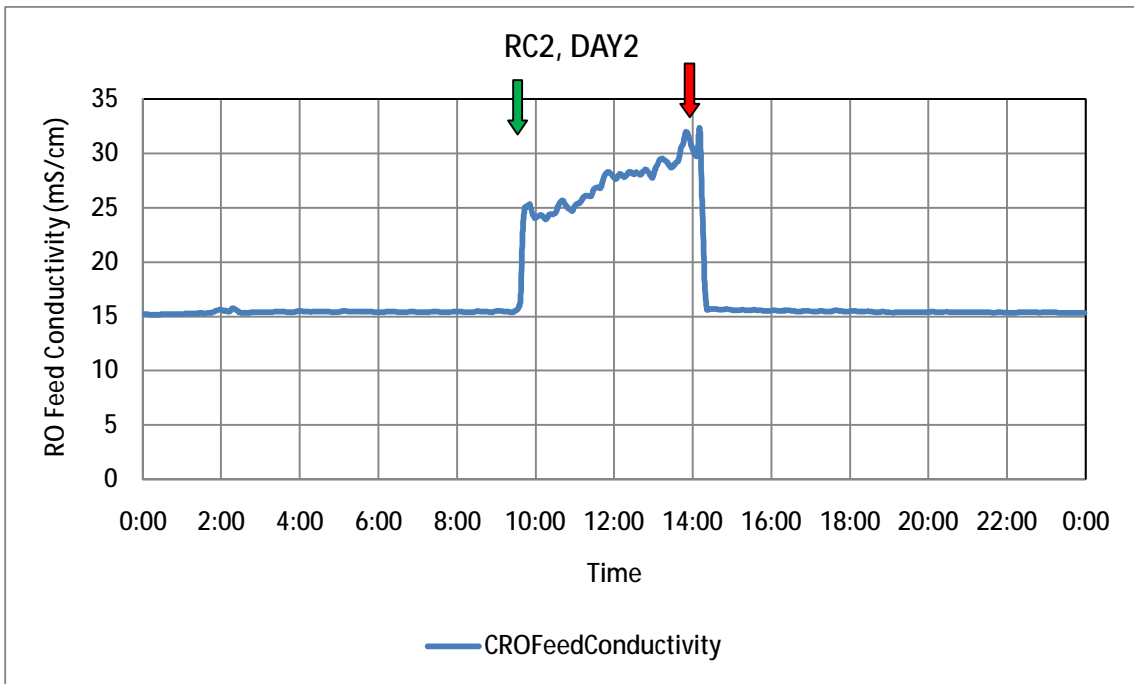


Figure 3.3.1.2.10: RO feed conductivity, RC2, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively

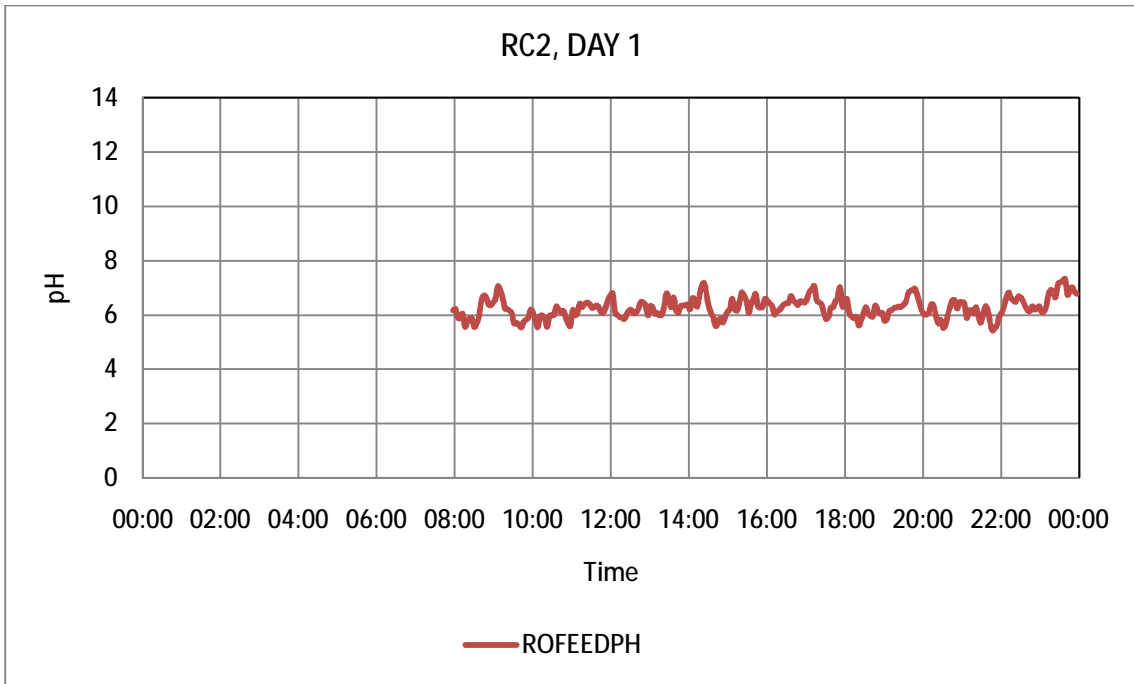


Figure 3.3.1.2.11: RO feed pH, RC2, Day 1

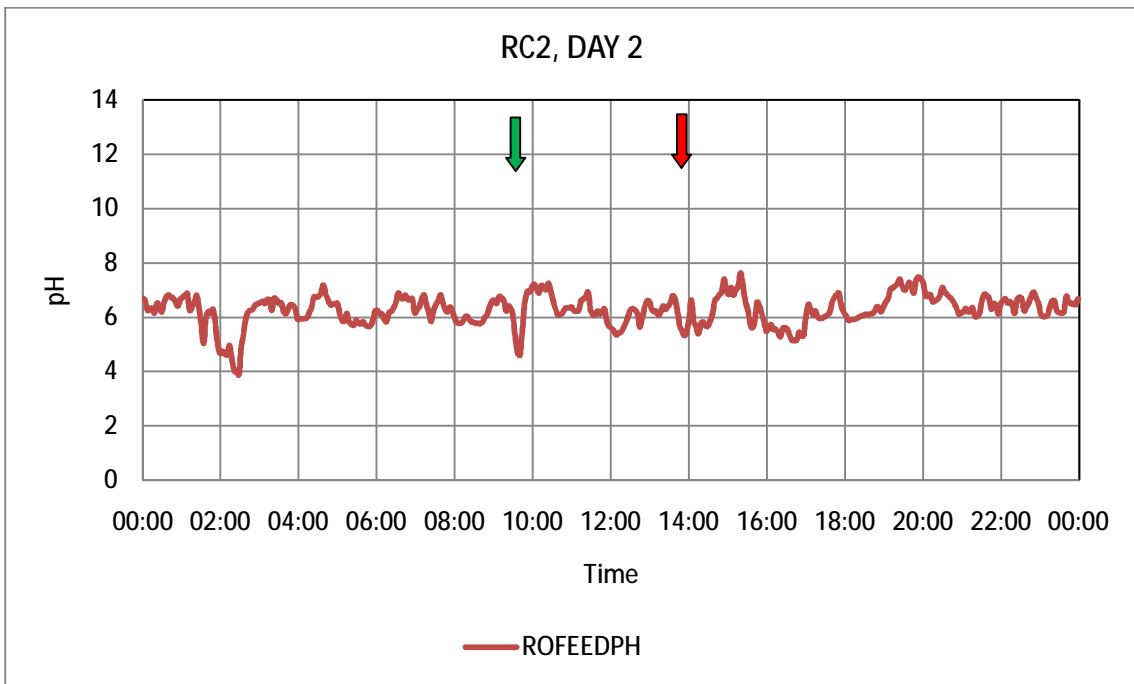


Figure 3.3.1.2.12: RO feed pH, RC2, Day 2, connection and disconnection of the CUF filtrate to/from the RO feed line are indicated by the green and red arrows respectively