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Report No. 195

AquaSel Technology Pilot-Scale Demonstration Menifee Desalter

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14. ABSTRACT
Over a nine-month period, between November 2014 and July 2015, EMWD tested GE's AquaSel technology on RO concentrate from the Menifee groundwater desalter. The test work was undertaken in three separate phases. Phases 1 and 2 were focused on troubleshooting different aspects of the process and optimizing the performance. Phase 3 focused on achieving up to 1,000 hours of operation at one set of operating conditions in the AquaSel and with the Menifee Desalter operating at 75 percent recovery. A total of 962 operating hours were achieved during Phase 3, and throughout that period the AquaSel system operated at 80 percent recovery, resulting in an overall recovery of 95 percent for the system.

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Prepared for the Bureau of Reclamation under
Agreement No. R14AP00171

by

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The mission of the Bureau of Reclamation is to manage, develop, and protect water and related resources in an environmentally and economically sound manner in the interest of the American public.

Disclaimer

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ACRONYMS AND ABBREVIATIONS

AF	acre-foot
AFY	acre-feet per year
BW	backwash
CIP	clean-in-place
DC	direct current
ED	electrodialysis
EDR	electrodialysis reversal
EMWD	Eastern Municipal Water District
GE	GE Water & Process Technologies
gpm	gallons per minute
lb/hr	pounds per hour
mgd	million gallons per day
mg/L	milligrams per liter
O&M	operation and maintenance
RO	reverse osmosis
TDS	total dissolved solids
µS	microsiemens

GLOSSARY

Anion. A negatively charged ion resulting from the dissociation of salts, minerals, or acids in water.

Anion (or anionic) membrane. A membrane through which only anions will transfer.

Anode (positive electrode). The electrode that attracts negatively charged anions.

Calcium sulfate ($CaSO_4$) saturation. The point beyond which any further addition of $CaSO_4$ in a given solution will cause precipitation.

Candle filter. A back washable, batch operated filter with candle-shaped filter elements arranged vertically inside a pressure vessel, operated as outside-in filtration.

Cathode (negative electrode). The electrode that attracts positively charged cations.

Cation. A positively charged ion resulting from the dissociation of salts, minerals or acids in water.

Cation (or cationic) membrane. A membrane through which only cations will transfer.

Cell pair. Repetitive section of a membrane stack consisting of a cation membrane, a demineralized water-flow spacer, an anion membrane and a concentrate water-flow spacer.

Concentrate stream. The stream in the membrane stack into which ions are transferred and concentrated.

Conductivity. The ability of a solution to conduct electrical current, commonly expressed in microsiemens/centimeter (micromhos/cm).

Cross leakage. Refers to the water leakage between demineralized and concentrate streams in the membrane stack.

Demineralize (desalt). To reduce the quantity of minerals or salts in an aqueous solution.

Product (demineralized, dilute) stream. The stream in the membrane stack from which ions are removed.

Electrodialysis (ED). A process in which ions are transferred through membranes from a less concentrated to a more concentrated solution as a result of the passage of a direct electric current.

Electrodialysis reversal (EDR). An electrodialysis process in which the polarity of the electrodes is reversed on a prescribed time cycle thus reversing the direction of ion movement in a membrane stack.

Electrical staging. The addition of electrode pairs in ED/EDR systems to optimize the DC electrical system within a membrane stack.

Electrode. A thin metal plate which carries electric current into and out of a membrane stack, normally constructed of platinum coated titanium alloys.

Electrode compartment (stream). The water flow compartment containing the metal electrode where oxidation/reduction reactions occur.

Fouling. A phenomenon in which organic or other materials are deposited on the membrane surface, causing inefficiencies.

Heavy cation membrane. A cation membrane made twice normal thickness (1.0 mm) to withstand greater differential pressures.

Hydraulic staging. Multiple passes of water between electrodes used in ED/EDR systems to achieve further demineralization.

Membrane (ED). An ion exchange resin cast in a sheet form which is essentially water-tight and electrically conductive.

Off-spec product. Product water that does not meet purity specifications.

One-pass. See candle filter.

Percent recovery. The percentage of feed water that becomes product water (the amount of product water produced divided by the total amount of feed water, multiplied by 100).

Precipitation. The action or process whereby solid particles form within a solution.

Scaling. The formation of a precipitate on a surface in contact with water as the result of a physical or chemical change.

Water flow spacer. A die-cut sheet of plastic which forms discrete flow paths for the demineralized and concentrate streams within an ED membrane stack.

Water transfer. Phenomenon in which water molecules are transferred through a membrane along with an ion.

WATSYS. GE design software used to calculate EDR process parameters. Inputs include feed water analysis, recovery, and product rate. Outputs include stack voltages and currents, stream flows and compositions, scaling potential, pumping and electrical power requirements.

METRIC CONVERSIONS

The metric equivalents for non-metric units used in the text are as follows:

Unit	Metric equivalent
1 acre-foot	1,233 cubic meters
1 foot	0.305 meters
1 gallon	3.79 liters
1 gallon per minute	3.79 liters per minute
1 kilowatt-hour per thousand gallons	0.264 kilowatt-hours per cubic meter
1 million gallons per day	3,785 cubic meters per day
1 pound	0.45 kilograms
1 pound per square inch	6.9 kilopascals
1 ton	907 kilograms

1.0 EXECUTIVE SUMMARY

In order to increase the supply of usable water in the United States, technologies focused on increasing recovery and decreasing waste from the treatment of impaired water sources need to be developed. To help achieve these goals, the AquaSel process aims at increasing desalter recovery and reducing brine waste streams from reverse osmosis (RO) processes.

The AquaSel process combines electrodialysis reversal (EDR) technology with a precipitation step added to the concentrate recycle loop. One of the limitations of EDR technologies is scaling in the concentrate loop. Typically, the EDR process can only recover water up to the point that the solubility limits of the sparingly soluble salts in the concentrate loop are exceeded. Adding the precipitation step is intended to remove calcium sulfate from the EDR process in a controlled environment, preventing precipitation in the membrane stack and increasing the recovery of the process.

Over a 9-month period, the AquaSel process was tested at the Menifee and Perris Desalter facility of Eastern Municipal Water District (EMWD), at Menifee, California. The primary piloting objective was to evaluate the AquaSel system as a brine reduction treatment option and to increase the desalter recovery from 75 to 95 percent, while producing an acceptable quality product stream.

Both of these goals were achieved and demonstrated during the testing periods. During the final phase of testing, the Menifee RO recovery was set to 75 percent, and the AquaSel recovery was set to 80 percent, resulting in an overall recovery of 95 percent. The 8-gallons-per-minute (gpm) pilot system was operated for 962 hours under the same conditions and was able to produce product water comparable to the raw well water that is used for blending with treated water from the desalters. Solids did precipitate during the pilot study, as shown by the increasing solids concentration in the precipitator, and so excess solids were sent to waste. The testing did reveal some shortcomings of the precipitator arrangement that would need to be addressed in any subsequent work.

A preliminary cost estimate showed that using the AquaSel system would make economic sense where current brine disposal costs are high and where the cost of alternative water sources is also high.

The AquaSel system shows promise as an approach to treat brine streams to achieve near-zero liquid discharge. Based on the data collected during the pilot study, the next step would be to proceed with a 100-gpm demonstration-scale system, in which design changes to the precipitation step can be implemented and the technology can be proven at a larger scale.

2.0 BACKGROUND

As water scarcity becomes more of an issue in many regions throughout the United States, there is a growing interest in desalination of impaired water sources. One of the major limitations of desalination is the concentrated waste stream that is produced by traditional technologies such as reverse osmosis (RO). Typically, RO can recover between 70 and 85 percent of the influent water from brackish sources, depending on the chemistry of the feedwater, resulting in a significant amount of concentrate that requires disposal. For brackish sources in which sodium chloride is the predominant solute, recovery levels of 90 percent are possible. However, these are not the focus of this study. The disposal of the concentrate stream is often challenging and can be cost prohibitive for locations where ocean disposal is not feasible. Even for inland regions of Southern California, where regional concentrate pipelines to the ocean are available, concentrate disposal is becoming more costly and more challenging due to issues with pipeline scaling, maintenance, and decreased line capacity. Figure 1 is a photograph of a portion of the Inland Empire Brine Line (formerly the Santa Ana Regional Interceptor line), showing internal scale formation.

Figure 1. Pipeline scaling of highly concentrated brine lines (Santa Ana Watershed Project Authority, 2010).

To reduce the cost of concentrate disposal, the recovery of the desalting process needs to be increased. However, increasing recovery can be challenging because the overall recovery of a desalination process is determined by the concentration of the least soluble of the sparingly soluble salts present (e.g., calcium carbonate, calcium sulfate, silica). To recover water beyond the solubility limit, solid salts must be removed from the process.

Eastern Municipal Water District (EMWD) owns and operates two groundwater desalters, the 3.1-million-gallons-per-day (mgd) Menifee Desalter and the 5.6-mgd Perris I Desalter, which together produce as much as 2.4 mgd of concentrate that is discharged to the Inland Empire Brine Line. Several processes, including lime softening followed by a secondary desalting unit, have been tested on the concentrate from the Menifee Desalter (Bureau of Reclamation, 2008). While these processes successfully reduce concentrations of sparingly soluble salts, they can use a significant amount of chemicals and produce a large amount of solid waste.

To reduce the amount of chemical used and waste produced, the AquaSel system couples a precipitation step with conventional EDR technology. This technology approaches concentrate minimization from a different angle by allowing salts to precipitate, in a controlled manner, in a precipitator. The approach makes use of EDR as a secondary desalting process by connecting it to the concentrate line of the existing RO process train, and uses a precipitation step within the concentrate loop to remove calcium sulfate before it can scale the EDR membrane stack. The precipitator receives a supersaturated solution from the EDR concentrate loop and allows it to achieve normal saturation by precipitating salts naturally.

2.1 Description of Unit Processes

2.1.1 Electrodialysis Reversal

EDR is an electrochemical separation process that uses a direct current (DC) voltage and ion exchange membranes to desalinate water. A schematic indicating the principal of electrodialysis is shown in Figure 2. A cathode and anode are separated in a sodium chloride solution by cationic and anionic membranes. When a direct current charge is applied to the electrodes, the ions begin to migrate to the oppositely charged electrode. However, since cationic membranes only allow passage of positively charged ions, and anionic membranes only allow passage of negatively charged ions, an ion-free (product) compartment and a concentrate compartment are created. The cationic and anionic membranes can be arranged in a stack configuration to produce product and concentrate streams from the incoming feed water.

To achieve a high rate of recovery, most of the concentrate leaving the stack is recycled, and a small amount is sent to waste to prevent salts from precipitating or scaling the membranes due to excessive concentrations. Wasted flows are made up with feed water. Figure 3 depicts a typical EDR process flow diagram.

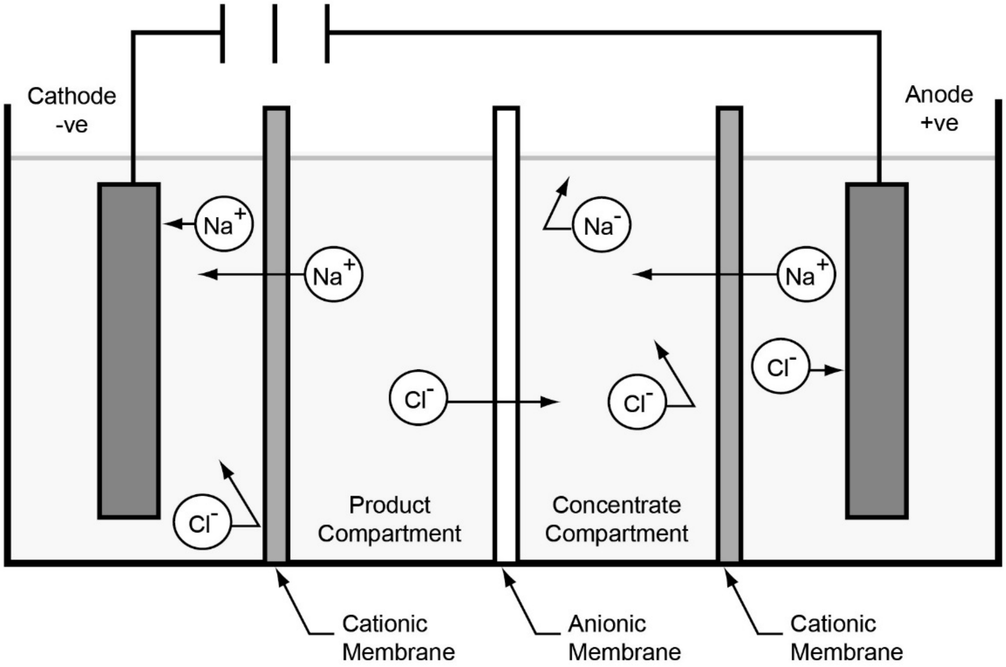


Figure 2. Principal of the electrodialysis process.

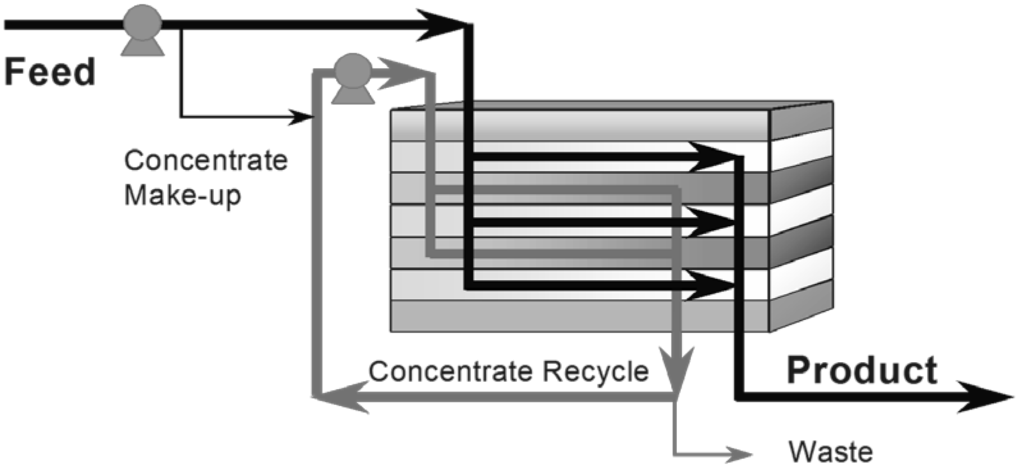


Figure 3. EDR process flow streams.

In the EDR process, the electrical polarity, and thus the demineralized and concentrate flow passages are automatically “reversed” two to four times every hour. This reverses the direction of ion movement, which provides “electrical flushing” of scale-forming ions and colloidal matter from the membrane surfaces, eliminating the need for extensive pretreatment of feed water and reducing chemical cleaning procedures. The effect of this polarity reversal is that the concentrate stream remaining in the stack, whose salinity is higher than the feed water, must be desalted. For a brief period after polarity reversal, the resulting product stream has an unacceptably high salinity and is known as “off-spec product." Conductivity-controlled valves shunt the product stream to waste until its salinity declines to an acceptable level.

2.1.2 AquaSel

In order to improve upon the standard EDR technology, GE Water & Process Technologies (GE) has added a precipitation step to the brine loop, creating what they refer to as the AquaSel system. This precipitation step focuses on removing calcium sulfate from the brine system in a controlled environment and thus prevents precipitation within the membrane stack. Instead of recycling directly back to the concentrate side of the EDR stack, the concentrate stream is fed into a precipitator.

The precipitator has two main sections. The interior column of the precipitator is the active mixing zone. Concentrate is fed to the interior column, continuously recirculating the high salinity water to control precipitation and particle size. The outer section operates as a standard upflow clarifier. Water flows slowly and gently upwards, allowing calcium sulfate particles to settle out of suspension while clearer water flows over the top and out to a filtration step. For this pilot study, the filtration step includes two options: candle filters or cartridge filters. The candle filters are designed to allow backwashing to automatically clean any particles off the interior surface on a regular basis. Cartridge filtration was also available as a backup to the candle filters and could be selected through manual isolation valves. Figure 4 shows the simplified AquaSel process flow diagram.

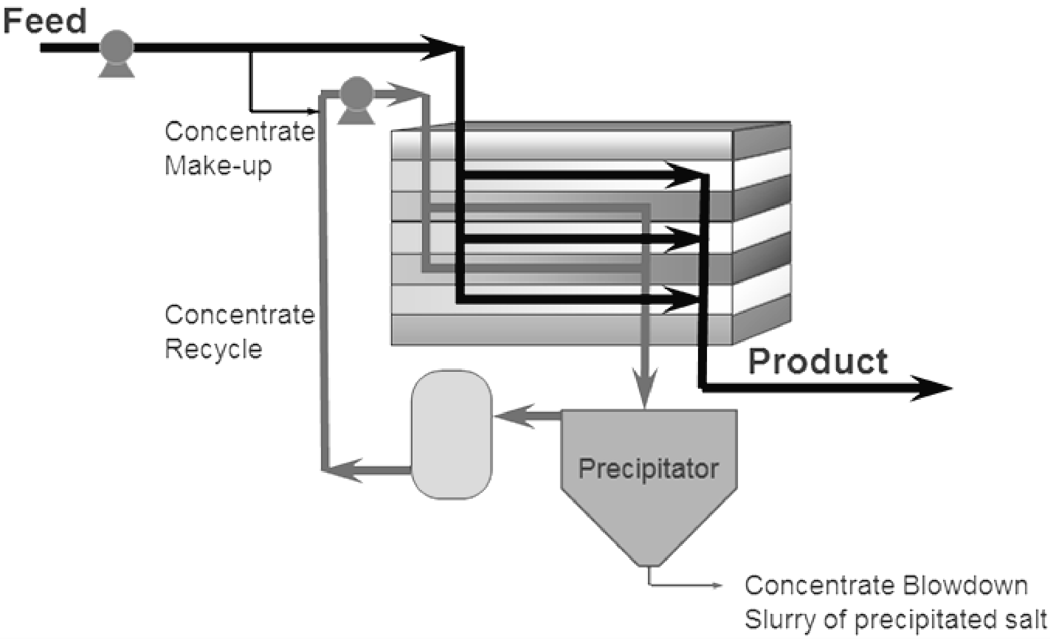


Figure 4. Simplified AquaSel process flow diagram.

2.2 Project Goals and Objectives

The overall purpose of this pilot study is to determine the feasibility of secondary treatment of concentrate from the Menifee and Perris I Desalters to reduce the volume of brine for final disposal and increase the overall recovery of potable water to around 95 percent or greater.

The pilot plant objectives include:

1. Obtain consistent, steady state operation.
2. Demonstrate the technical feasibility of the AquaSel process operating on Menifee RO concentrate by achieving about 1,000 hours of continuous operation at one set of operating conditions defined in Table 1.

Table 1. AquaSel Operating Conditions for Final Testing Phase

Parameter	Value
Menifee Desalter recovery, percent	75
Raw feed water flow rate, gpm	8
AquaSel system recovery, percent	80
EDR blowdown permeate conductivity, µS/cm	<3,200
Solids blowdown rate, gallons per hour	6
Precipitator solids concentration, percent (by volume)	11–12
Precipitator solids concentration, percent (by weight)	5–6

3. Demonstrate stable performance of the AquaSel system through continuous monitoring of the EDR system pressures, EDR system voltages and amperages, and feed, product, and concentrate conductivities, as well as the performance of the precipitation tank, and consistent product water quality that could be returned to EMWD's potable water system.
4. Determine the solids production rate from the process and demonstrate the solids mass balance for the system.
5. Determine the chemical make-up of the solids produced by the process and ascertain disposal options for solids from a full-scale facility treating 2.4 mgd of primary desalter concentrate.
6. By monitoring and reporting the cost for the pilot system, determine the anticipated long-term operating costs for the 2.4-mgd system, including the dosage of all chemicals, electrical power costs, membrane cleaning frequency and costs, solids production rate and disposal cost, membrane and other EDR stack component replacement costs.

3.0 TECHNICAL APPROACH

This pilot study initially was intended to consist of a single phase in which the operations and performance of the AquaSel system would be measured and evaluated. However, the study was ultimately broken into three phases as issues arose and were resolved. The following sections describe the pilot plant facility and outline the major time periods for problem solving issues with the pilot.

3.1 Pilot Study Facility

A layout of the pilot testing equipment is shown in Figure 5. A detailed process flow diagram showing the configuration of the AquaSel during the final testing phase is shown in Figure 6. The system included three separate items: (1) a 40-foot shipping container that accommodated the EDR stack, controls, electrical systems, filters, and tanks; (2) a precipitator tank located just to the north of the shipping container; and (3) a roll-off container to store solids blown down from the process. The testing equipment was temporarily plumbed into the Menifee Desalter concentrate line, which supplied approximately 8 gpm to the AquaSel unit.

3.1.1 Source Water for Pilot Testing

The Menifee Desalter is an existing groundwater RO facility owned and operated by EMWD. This 3.1-mgd facility includes preliminary filtration of the RO feed water through 5-micron cartridge filters before it enters the RO treatment process. The RO plant consists of two treatment trains, each in a two-stage array and operating at 75-percent recovery. Table 2 presents typical concentrations of constituents in the Menifee Desalter concentrate.

Table 2. Menifee Desalter Concentrate – EDR Feedwater Quality[1]

Parameter	Units	Value
Alkalinity as $CaCO_3$	mg/L	4,075
Total dissolved solids (TDS)	mg/L	7,965
Sulfate	mg/L	666
Sodium	mg/L	1,159
Calcium	mg/L	1,150
Magnesium	mg/L	293
Chloride	mg/L	3,581
Total silica	mg/L	166

[1] Menifee Desalter was operated at a recovery of 75 percent.

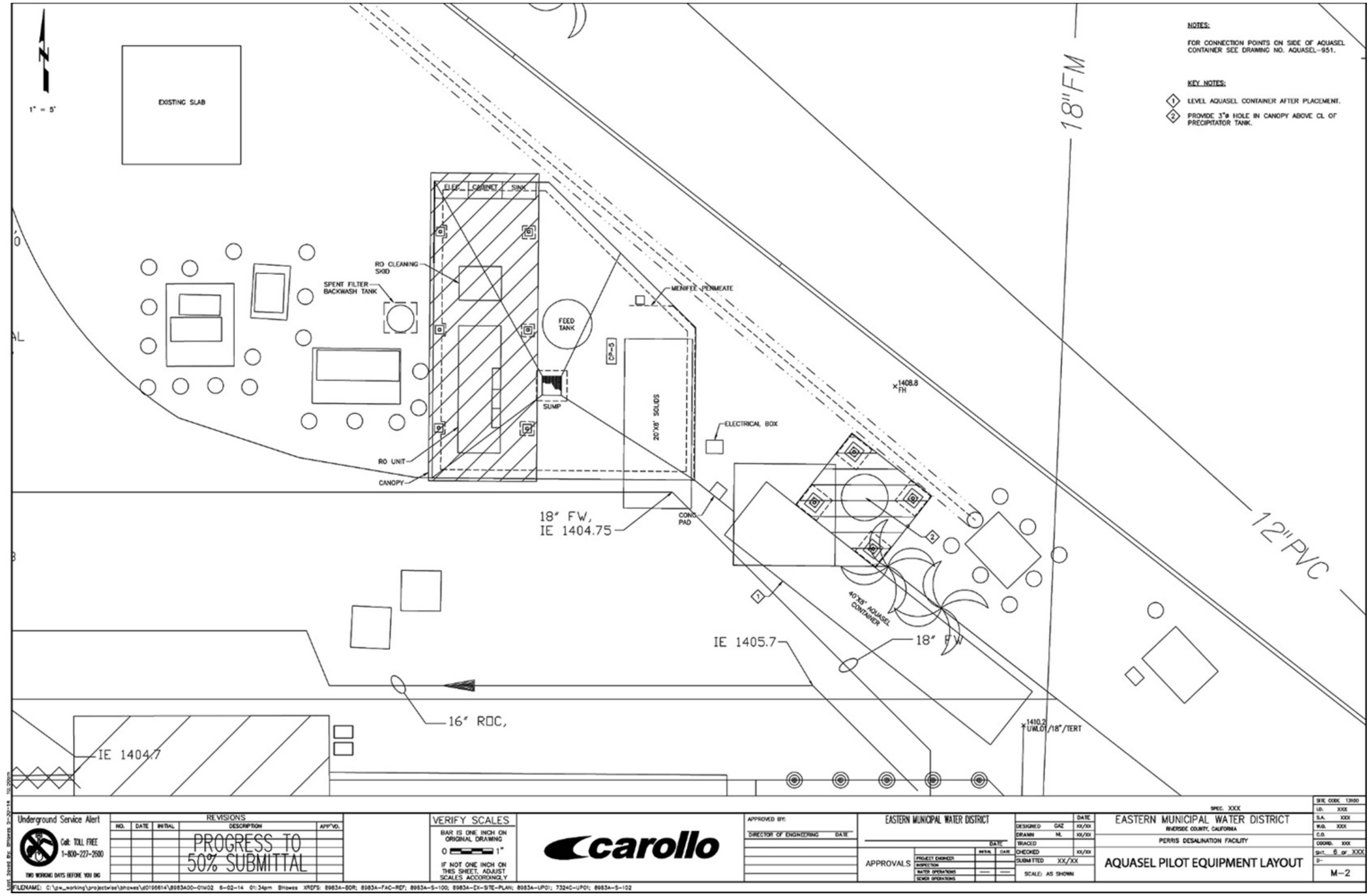


Figure 5. AquaSel pilot equipment layout.

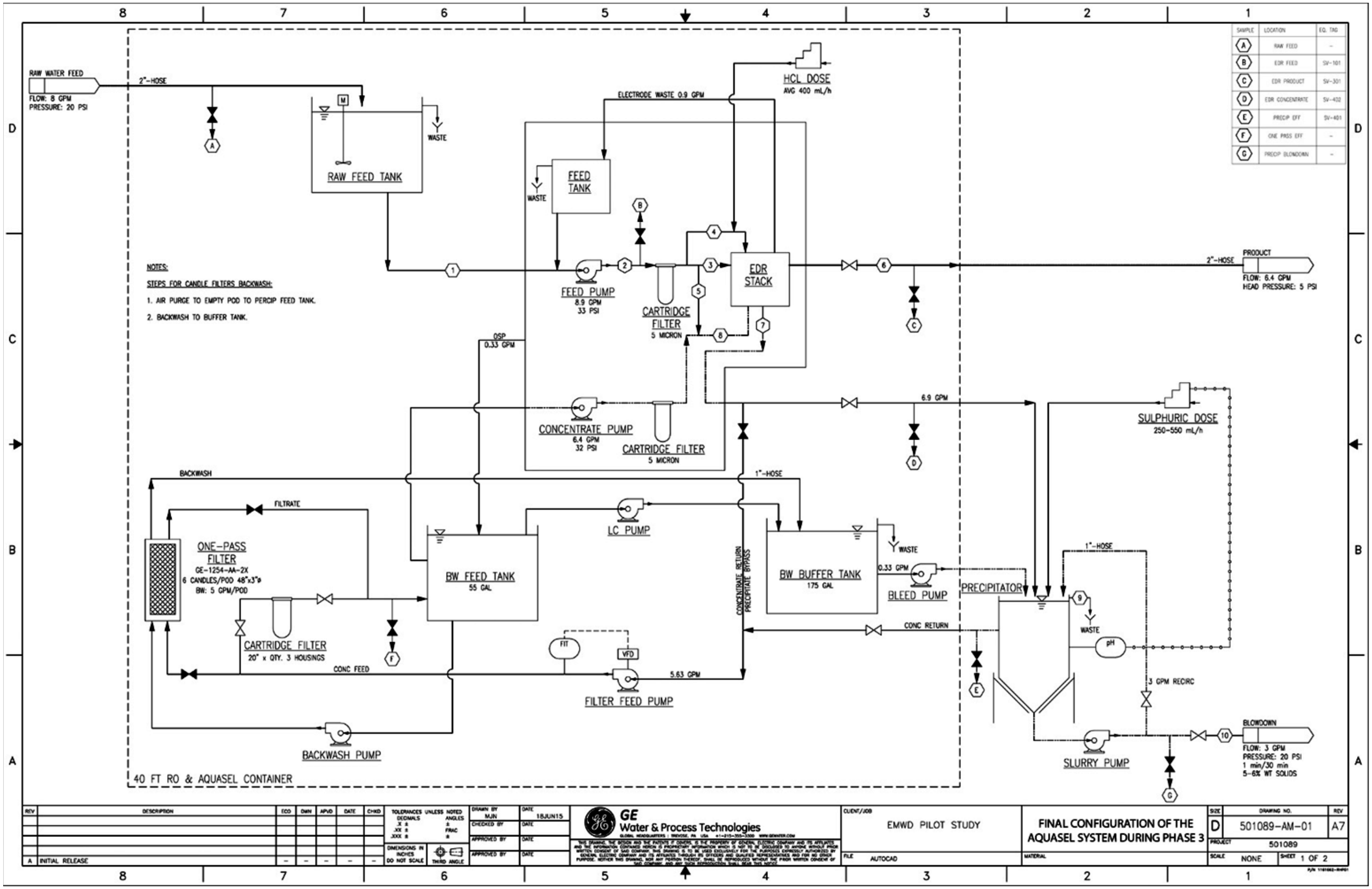


Figure 6. Final configuration of the AquaSel system during Phase 3.

3.1.2 AquaSel Design Criteria

GE supplied the AquaSel pilot unit, which consisted of one membrane stack with two electrical and four hydraulic stages (50 cell pairs per hydraulic stage). An electrical stage comprises one cathode and one anode separated by a series of cationic and anionic membranes and spacers. The number of hydraulic stages indicates the number of passes the product water makes through the stack. Specific attributes of the GE pilot EDR unit and the precipitator are summarized in Table 3.

Table 3. AquaSel Pilot Design Criteria

Parameter	Value
EDR process unit:	
Feed flow, gpm	8.0 [1]
Product flow, gpm	6.4
Blowdown flow, gpm	1.6 [2]
Brine makeup flow, gpm	0.9
Overall recovery (percent)	80
Electrical stages	2
Hydraulic stages	4
Cell pairs	50/50/50/50
Power settings (volts, amps)	Stage 1 – 47V, 35A Stage 2 – 35V, 19A
Reverse polarity setting (min)	15
Precipitator process unit:	
Hydraulic retention time (min)	120
Solid concentration (percent)	11–14 (by volume) 5–6 (by weight)
Solids blowdown rate, gal per hour	6

[1] To match the mass balance.

[2] 1.5 gpm of concentrate blowdown plus 0.1 gpm of precipitator blowdown.

RO brine water was fed into a raw feed tank, essentially an equalization tank, and then was pumped through a 5-micron cartridge filter before entering the dilute stream of the EDR stack. The dilute stream was desalted in the EDR stack and discharged as product water. A small stream of feed water was diverted to flow through the electrode streams in the EDR stack and initially included hydrochloric acid (HCl) to adjust the pH. Electrode streams were sent to waste. Another portion of feed water served as the make-up stream for the brine loop.

The concentrate line was continually recirculated. Starting at the EDR, the concentrate stream was pumped through the EDR stack and was further concentrated with ions removed from the dilute stream. From the EDR stack, the concentrate was fed directly to the precipitator's inner ring. This ring was continuously mixed with two impeller blades, resulting in a concentrated slurry of calcium sulfate particles. The concentrate stream then flowed through the remaining sections of the precipitator like a standard upflow clarifier. Sulfuric addition was used to maintain the pH at 6.3.

To start the process, calcium sulfate dehydrate (gypsum) was seeded to achieve a 10 percent by volume solids concentration in the center ring of the precipitator. Solids were wasted from the precipitator using the recirculation stream by sending slurry to a waste bin. Additionally, some clarified concentrate overflowed and went to waste. The remaining clarified concentrate was pumped using the filter feed pump through a filtration step.

The filtration step included two options: a candle filter system or a series of three parallel 5-micron cartridge filters. This step was designed to remove any remaining solids before the recirculation flows were sent back through the EDR. The default operating condition was the candle filter, which could be backwashed. The cartridge filters, which must be replaced after use, were provided as a backup system.

After the filtration step, the concentrate was collected in the backwash (BW) feed tank and blended with off-spec product water generated by polarity reversal of the EDR stack. Concentrate from this tank was pumped back through the EDR stack to complete the concentrate loop. Excess water from the BW feed tank was pumped to the BW buffer tank and returned to the precipitator using a separate pump.

3.2 Pilot Plant Setup, Commissioning, and Operating Protocol

As mentioned, this study was intended to operate as a single testing phase. However, when the pilot began operation, unforeseen problems arose requiring dedicated troubleshooting. The troubleshooting efforts and modifications are discussed briefly in this section and in detail in the Discussion. The operation of the pilot substantially changed from what was originally intended. The following sections outline the major time periods for problem solving issues.

3.2.1 Phase 1: Pretreatment Review – November 2014 to March 2015

Shortly after the pilot first began operating, the EDR stack inlet pressure transmitters quickly showed significant pressure increases. Some investigation indicated substantial silica fouling of the pilot system. Phase 1 is characterized as the period of pilot operation focused on the pretreatment evaluation, aimed at minimizing silica scaling within the pilot process. A more lengthy discussion of what happened during Phase 1 can be found in Section 4.1 of this report.

3.2.2 Phase 2: Process Stabilization – March 2015 to May 2015

Once the silica issue was resolved, the next phase focused on problems preventing the pilot from operating at steady state. During this phase, the flow balance had to be attained and maintained to a satisfactory degree before final testing could begin. Unbalanced flows caused flow surges and solids carryover, leading to unexpected shutdowns and frequent alarms. Throughout the course of this phase, as flows were evened out, the process stabilized. Shutdowns became less frequent, and the system was able to run continuously for a few days without alarm. Figure 7 shows the pressure stabilization from March through May. A detailed discussion of what transpired during Phase 2 can be found in Section 4.2 of this report.

3.2.3 New Membrane Installation – April 28, 2015

Once multi-day operation was established and pressures stabilized (as shown in Figure 7), the anion membranes in the EDR stack were replaced. Now that the process was stabilized, it was important to show how a new unit would be expected to operate. With new membranes, there was a brief dip in operating pressure, but by early May the pressures returned to a range typical of stabilized operation. The goal of replacing these membranes was to improve the product water quality. Since the previous membranes had been fouled by silica, the product water did not meet the quality goals and could not be expected to. With the membranes replaced, the flows stabilized, and the scaling under control, the system was ready for rigorous testing.

3.2.4 Phase 3: Process Evaluation – June 2015 to July 2015

Once the process was fully stabilized and the anion membranes were replaced, the system was ready to begin preparations for final testing. Between early May and the middle of June, the system was monitored and inspected to ensure typical operating conditions for a full-scale system. Once final testing began, the intent was to evaluate potential operating conditions of a full-scale system and establish design parameters based on steady state conditions. This process was vital in determining the level of performance that could be expected of a large-scale installation. The final test period, known as Phase 3, officially began on June 19, 2015. A detailed discussion of the events in Phase 3 can be found in Section 4.3 of this report.

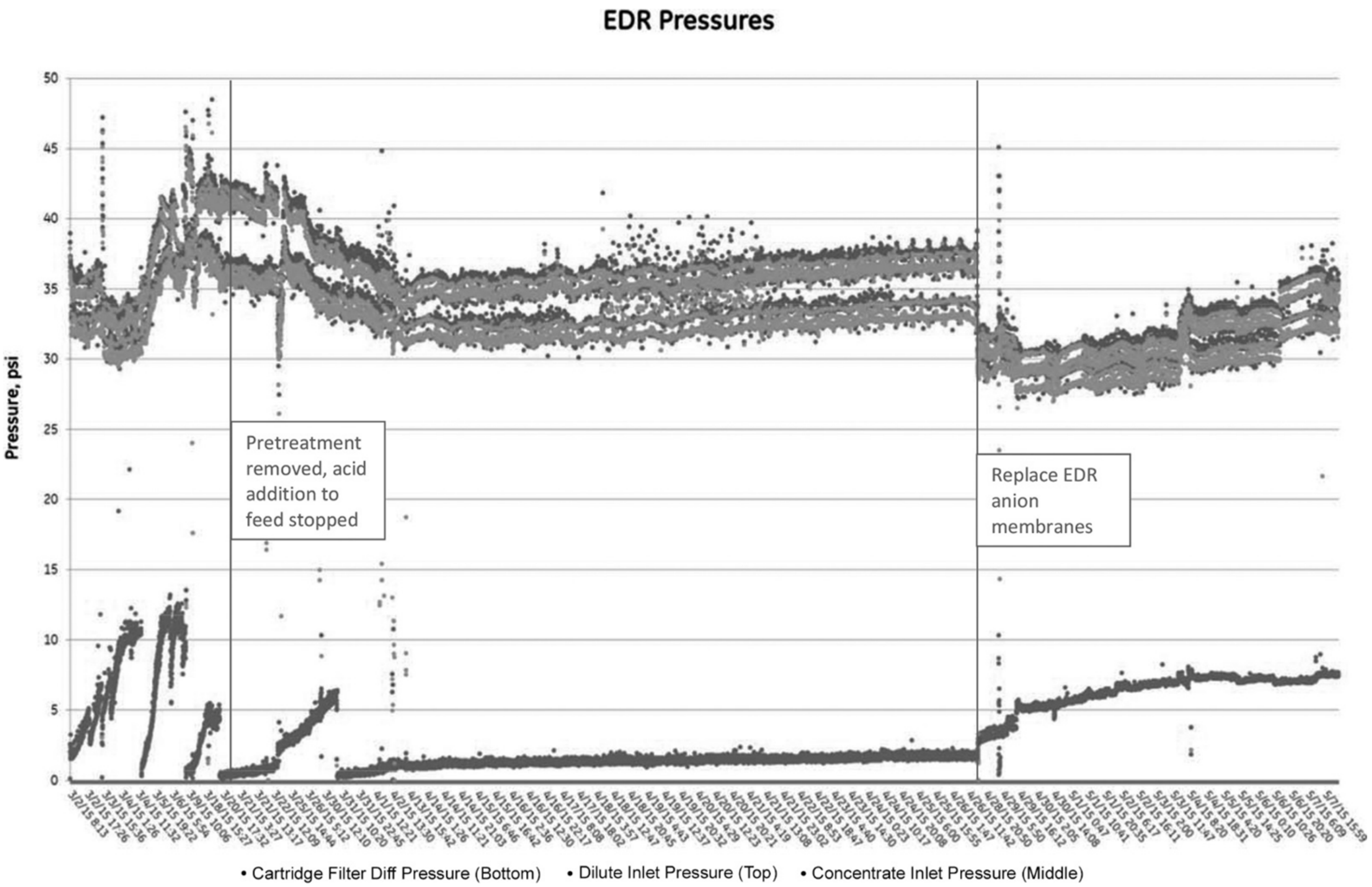


Figure 7. EDR Pressures between March 1 and May 8, 2015.

3.3 Sampling and Monitoring Protocol

Both manual and automated data collection systems were used during this pilot. Automated data collection included date, time, runtime, pH, conductivity, temperature, stack voltages, current drawn, pump speeds, electrode flows, concentrate recycle flows, concentrate blowdown flows, concentrate makeup flows, pressures, and differential pressures.

Grab samples from six points in the pilot system were analyzed by EMWD's certified laboratory. Table 4 lists the parameters analyzed and the number of tested samples from each location. Figure 8 shows a schematic of sampling locations used during the pilot testing.

3.4 Interpretation of Performance Data

EDR data is not normalized because there is no established normalization procedure. EDR analysis is usually conducted on hydraulic and electrical performance data. Hydraulic performance data is used to determine salt rejection, production, and recovery. Electrical data collected is used to determine the energy demand of the system at different recoveries and to generate a profile of the voltage, amperage, and resistance of each stage.

3.4.1 Salt Rejection

Salt rejection is an important parameter of EDR performance. In order to calculate salt rejection, the actual salt passage must first be calculated as the ratio of salt in the product stream divided by the salt concentration in the feed stream. The salt rejection, expressed as percent rejection, would therefore be equal to 100 percent minus the actual salt passage. For this study, because conductivity analyzers provide substantially more data than the occasional grab sample, salt concentrations are expressed in terms of TDS and calculated using empirical factors correlating TDS to conductivity. For feed streams, the empirical factor of 0.9 TDS/conductivity was used. Similarly for product streams, the empirical factor of 0.78 TDS/conductivity was used.

Table 4. Process Related Analysis

Parameter	Total Est. Lab Samples *(per 7-week testing phase)*	Type	Sampling Frequency (per week)					
			Stream No. 1: Menifee Brine	**Stream No. 2: EDR Feed**	**Stream No. 3: EDR Product**	**Stream No. 4: Precip Feed [1]**	**Stream No. 5: Precip Effluent**	**Stream No. 6: Solids Wasted**
Cartridge filter delta P [3]		L		10				
pH [3]		C/G [3,4]	3	3	3	3	3	3
Temperature [3]		C	-	-	-	-	-	-
Conductivity [3]		C/G	3	3	3	3	3	
Turbidity	105	G [4]	3	3	3	3	3	
Alkalinity	105	G	3	3	3	3	3	
Total dissolved solids (TDS)	126	G	3	3	3	3	3	3
Total organic carbon	126	G	3	3	3	3	3	3
Total suspended solids	126	G	3	3	3	3	3	3
Nitrate	105	G	3	3	3	3	3	3
Bicarbonate	105	G	3	3	3	3	3	
Total hardness	105	G	3	3	3	3	3	
Barium	105	G	3	3	3	3	3	
Strontium	105	G	3	3	3	3	3	
Sulfate	126	G	3	3	3	3	3	3
Sodium	126	G	3	3	3	3	3	3
Ammonia-N	35	G	3	3	3	3	3	
Potassium	35	G	3	3	3	3	3	
Total phosphorous	35	G	3	3	3	3	3	
Calcium	126	G	3	3	3	3	3	3
Total iron	105	G	3	3	3	3	3	
Ferric iron ($Fe^{3}+$)	105	G	3	3	3	3	3	
Magnesium	126	G	3	3	3	3	3	3

Parameter	Total Est. Lab Samples *(per 7-week testing phase)*	Type	Sampling Frequency (per week)					
			Stream No. 1: Menifee Brine	**Stream No. 2: EDR Feed**	**Stream No. 3: EDR Product**	**Stream No. 4: Precip Feed [1]**	**Stream No. 5: Precip Effluent**	**Stream No. 6: Solids Wasted**
Manganese	35	G	3	3	3	3	3	
Chloride	126	G	3	3	3	3	3	3
Fluoride	105	G	3	3	3	3	3	
Boron	105	G	3	3	3	3	3	
Aluminum	105	G	3	3		3	3	
Arsenic	105	G	3	3	3	3	3	
Copper	105	G	3	3		3	3	
Selenium	105	G	3	3	3	3	3	
Total silica	105	G	3	3	3	3	3	
Reactive silica	105	G	3	3	3	3	3	
Percent solids	21	G						3
Heterotrophic plate count	35	G	1	1	1	1	1	
Radon [5]	7							1
Gross alpha [5]	7							1
Solids analysis (percent $CaSO_4$; percent $CaCO_3$) [5]	7							1
Other - for potable quality permit (EMWD to provide)								

---

[1] Precipitator feed is the EDR brine stream.

[2] Measured onsite using portable equipment.

[3] C/L=Continuous data logging or logged by pilot plant operator.

[4] Sample type: G = grab.

[5] In precipitated solids. Analyzed by outside laboratory using scanning electron microscopy, energy dispersive x-ray spectroscopy, or x-ray photoelectron spectroscopy.

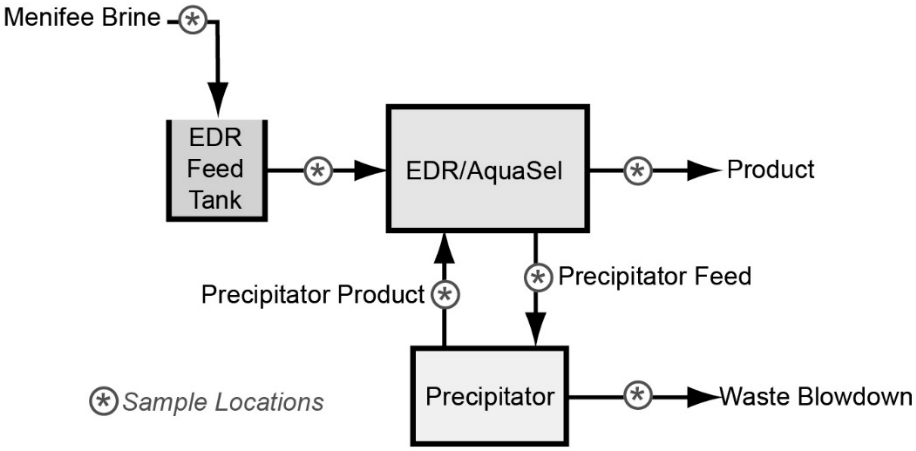


Figure 8. Schematic of AquaSel pilot system showing sample locations.

4.0 RESULTS AND DISCUSSION

The AquaSel pilot system was set up at EMWD's Menifee site during November 2014 and achieved its first full day of operation on November 30, 2014. Between December 2014 and July 2015, various performance tests were carried out on the unit.

4.1 Phase 1: Pretreatment Discussion

When the pilot began operation, the Menifee Desalter was operating at 75 percent recovery, and the AquaSel system was set to 90 percent recovery. After a short period of run time, the EDR stack inlet pressure transmitters quickly showed significant pressure increases. It was quickly evident that silica was building on both the cartridge filters and the EDR membranes with inlet silica concentrations in excess of the design recommended limit of 125 milligrams per liter (mg/L). The concentration of silica in the RO concentrate feeding the AquaSel was in excess of 140 mg/L. Figure 9 shows the EDR inlet dilute and concentrate pressures as well as the cartridge filter differential pressure.

The baseline pressure for the AquaSel system was approximately 30 pounds per square inch (psi). However, Figure 9 shows pressures in excess of 45 psi, or a 50-percent increase in operating pressure. After just 6 days of operation, an acid clean-in-place (CIP) operation was performed using a 5-percent solution of hydrochloric acid to remove the scale and return the stack to original operating pressures. However, as is evident from Figure 9, the pressures quickly climbed again.

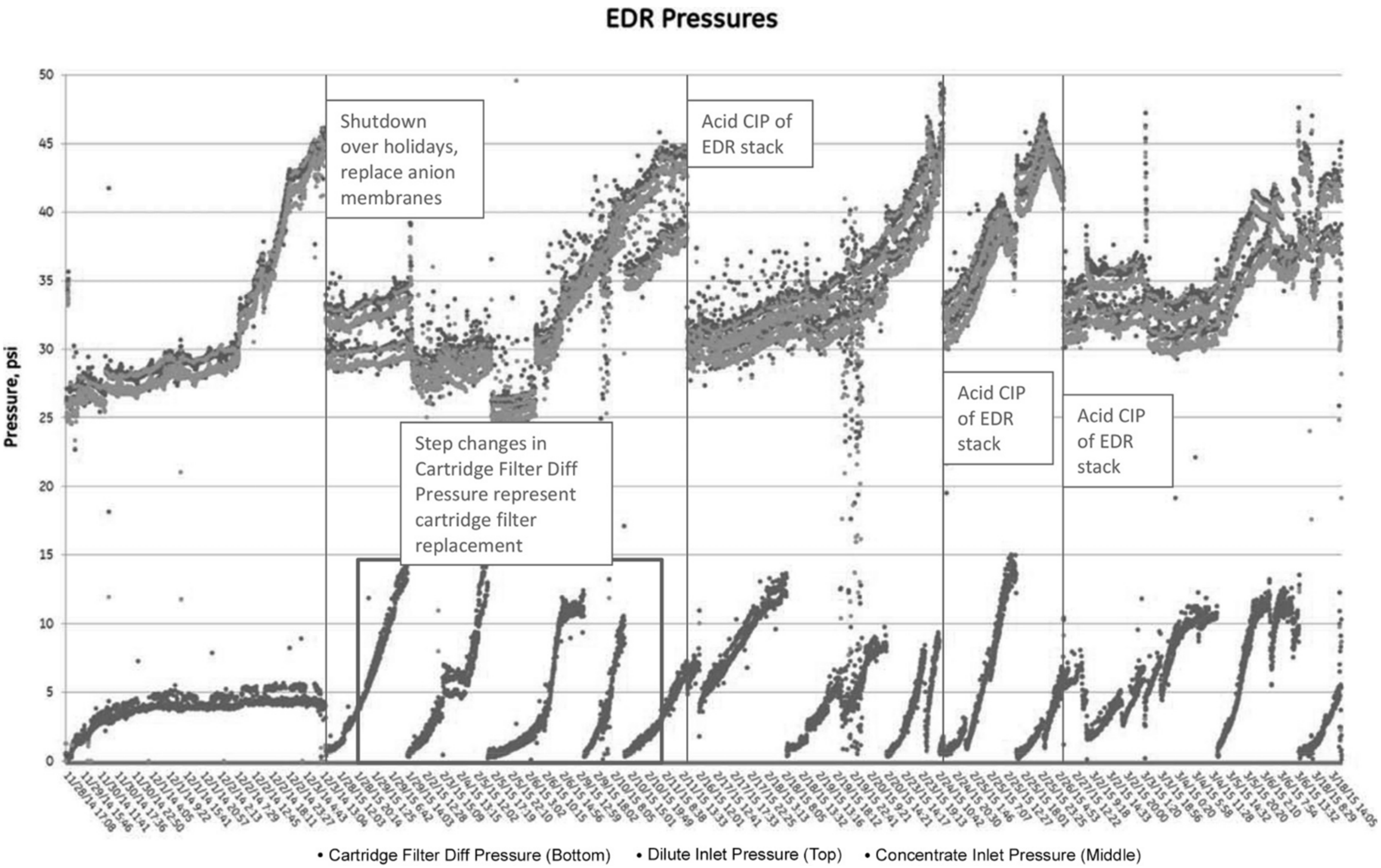


Figure 9. EDR pressure trend in Phase 1.

The cartridge filter assembly and the membrane stack were disassembled to further understand the problem. A coating of gelatinous material was covering the cartridge filter and membranes, as shown in Figure 10. The lab verified the material was in fact silica.

Figure 10. Silica-coated filter and membrane.

A multimedia filtration system with ferric chloride addition was incorporated into the process flow to address the silica fouling. This system included ferric chloride dosing to the mixed raw feed tank. The pilot feed pump would then push the RO concentrate through the multimedia filters, then through cartridge filters. Sulfuric acid was added to the cartridge filter effluent before it went to the EDR skid and the cartridge filters included in the shipping container. However, after a few weeks of operation with this in-line pretreatment system, there was no noticeable improvement to the system. The RO recovery upstream was decreased from 75 percent to 70 percent in an attempt to decrease silica concentration in the EDR feed, but the problem persisted.

In a previous pilot study, an EDR pilot had been successfully run treating similar RO concentrate without this same silica fouling issue. Sulfuric acid was continuously added to the feed of the AquaSel pilot. However, the previous pilot did not reduce EDR feed pH to 5.5, which was thought to be essential to control calcium carbonate scaling. The Menifee Desalter uses a King Lee antiscalant to keep silica in solution as part of standard operation prior to the AquaSel pilot. The GE process team contacted the manufacturer to better understand the occurring phenomenon.

King Lee informed GE that at a pH of 5.5, the antiscalant becomes inactive. As the chemical became less effective due to the low pH, the silica was able to drop out and deposit on the filters and membrane. Since another pilot had successfully run at a pH range of 7–8, a test was run to see if eliminating the acid addition would also eliminate the silica fouling. This test showed immediate success and thus, sulfuric acid dosing to the EDR feed was halted for the rest of the study.

However, substantial reduction of alkalinity was still required for EDR operation and was thus transferred to the concentrate side of the process. In order to achieve pH ranges between 5.5 and 6.0 there, additional acid dosing to the concentrate exceeded CO_2 solubility limits, initially leading to CO_2 gas generation in the precipitator and floatation of solids. After some fine-tuning, the optimal pH that did not exceed CO_2 solubility limits but still prevented calcium carbonate scaling was identified to be in a range of 6.3 to 6.5. Figure 11 shows pressure stabilization after the acid dose to EDR feed was eliminated and the pre-treatment system was bypassed.

4.1.1 Feed Water Quality

Table 5 shows the Menifee Desalter concentrate water quality during Phase 1.

Table 5. Menifee Desalter Concentrate – EDR Feedwater Quality[1]

Parameter	Units	Value
Alkalinity as $CaCO_3$	mg/L	4,075
Total Dissolved Solids (TDS)	mg/L	7,965
Sulfate	mg/L	666
Sodium	mg/L	1,159
Calcium	mg/L	1,150
Magnesium	mg/L	293
Chloride	mg/L	3,581
Total Silica	mg/L	166

[1] Menifee Desalter was operated at a recovery of 75 percent.

4.1.2 Hydraulic Performance

Process modifications were made during early operation to manage scaling in the front end of the system. Figure 12 illustrates the system pressures during Phase 1 operations testing. As shown, the pressures rapidly increased a few days after operation began due to fouling from scaling as the feed pumping maintained constant flow through the diluting side of the membrane stack.

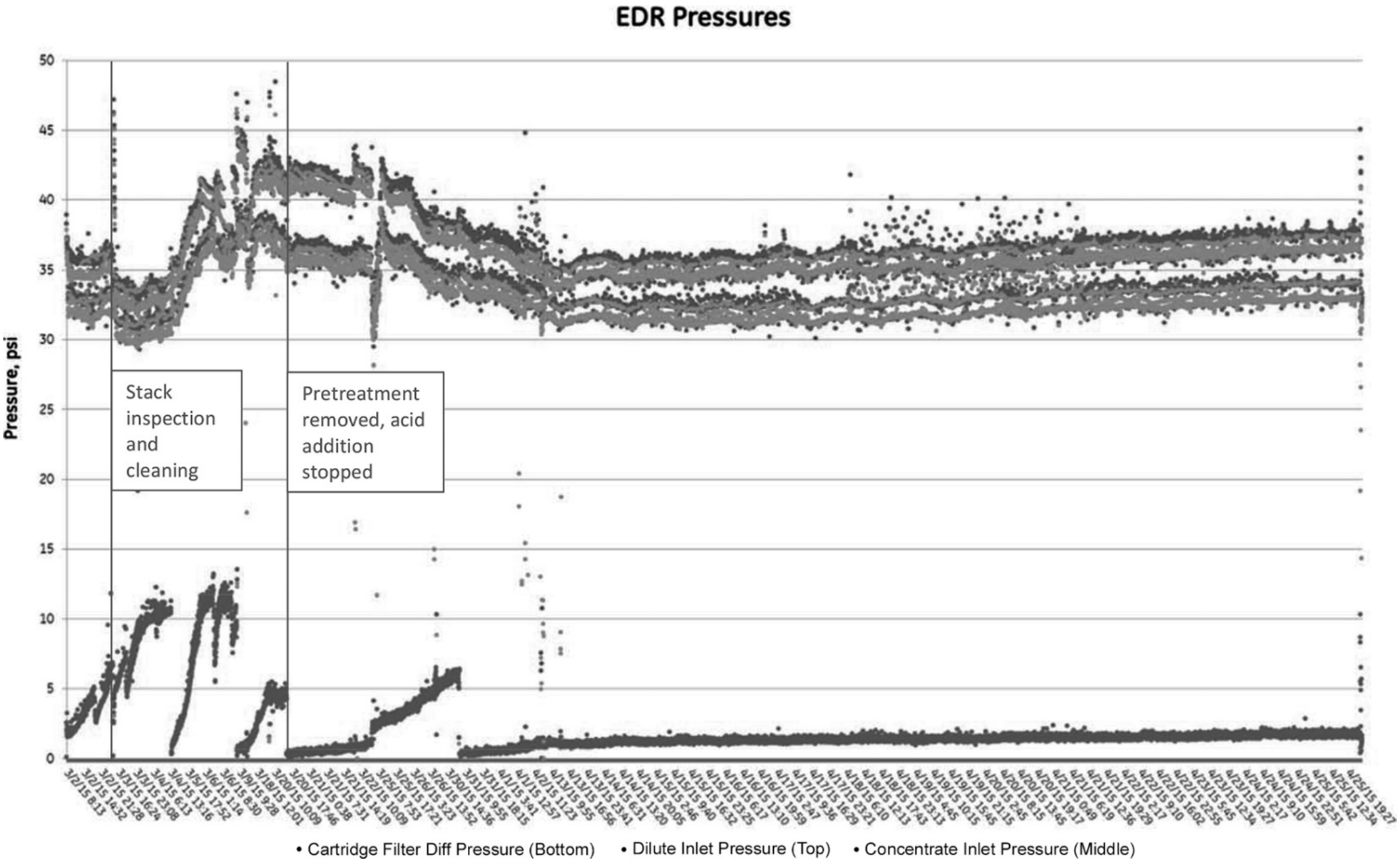


Figure 11. Phase 1 EDR pressures through April 26.

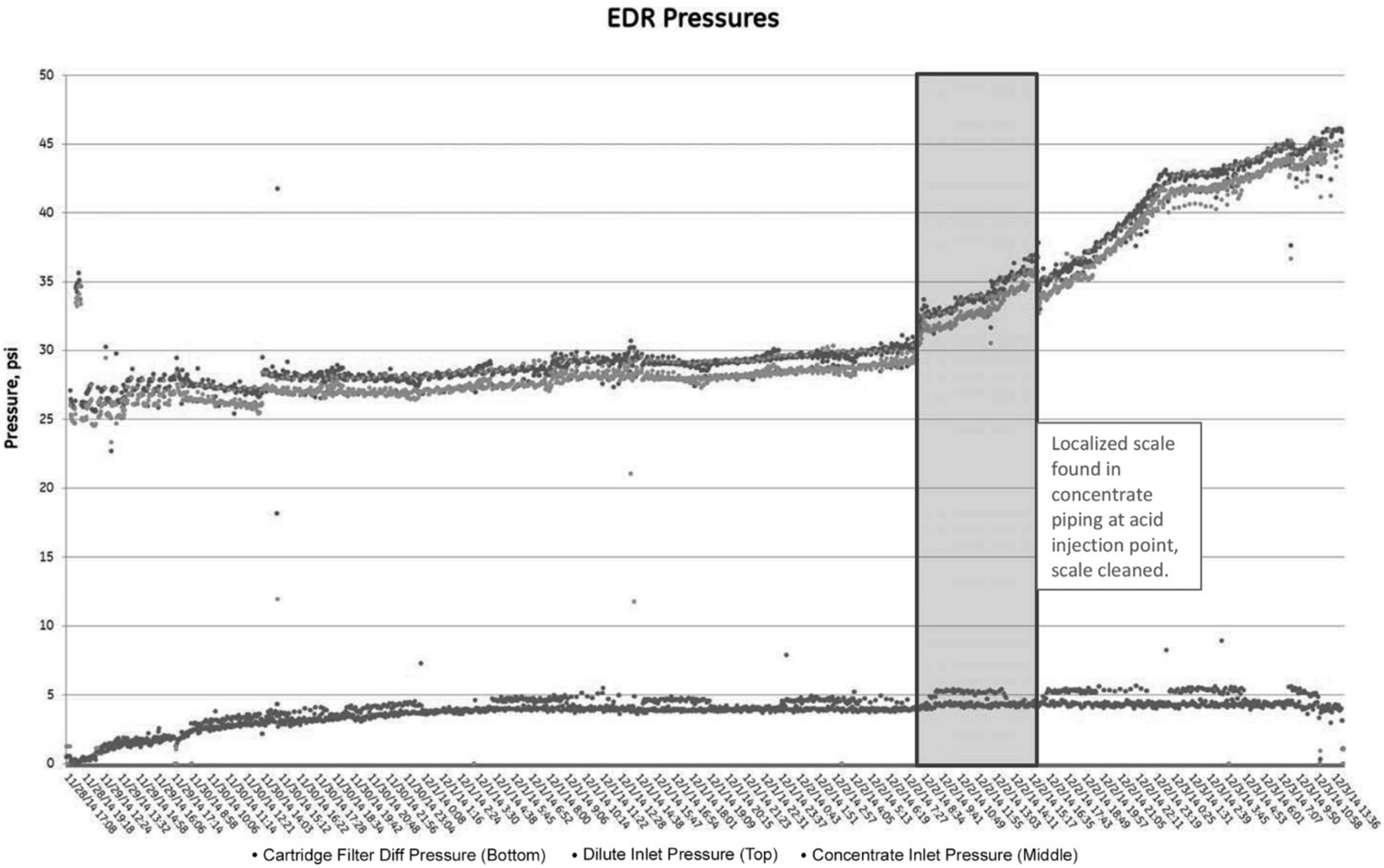


Figure 12. EDR pressures during Phase 1 operations.

4.1.3 Electric Performance

Figure 13 shows increasing Stage 2 amperages as membranes acclimatize and alter efficiency. However, Stage 1 and Stage 2 voltages remain fairly constant throughout pilot Phase 1 operation. Figure 14 shows a slight increase in stack power consumption during Phase 1 as the membranes acclimatize.

4.1.4 Water Quality

During this phase, product conductivity was better than expected (<3,000 µS/cm) and concentrate conductivity was effectively driven up to expected levels (>60,000 µS/cm). Table 6 shows data collected on site from November 30 to December 9 using onsite Hach test kits (for calcium and sulfate) and a handheld meter (conductivity and pH). It should be noted that these values differ from the Phase 1 values reported in Table 5 above, likely due to water quality at the time of testing and the test method.

Table 6. Operational Data – Phase 1

Parameter	**Units**	**Raw**	**Acidified Raw**	**Product**	**EDR Conc.**	**Precip. Effluent**	**One-Pass Effluent**
pH	units	7.51	5.60	5.50	6.06	6.56	-
Conductivity	µS/cm	10,780	10,730	2,154	67,650	56,740	-
Sulfate	mg/L	390	590	48	1,200	-	900
Calcium	mg/L	1,032	1,056	48	9,880	-	7,140

The above data supports the conductivity expectations for product and concentrate streams, and removal of salts in the precipitator ($CaSO_4$); however, the relative removals of calcium and sulfate through the precipitator were not consistent. The observed variability may have resulted from periodic instability created by the internal recycling of process streams, even though samples were taken at the same time. Figures 15 and 16 show conductivity plots representing Phase 1 of pilot testing.

Additionally, GE, using their WATSYS system, prepared a mass balance analysis for operations during Phase 1 (Figure 17). Figure 18 shows the simplified process flow diagram showing the Phase 1 operation schematic.

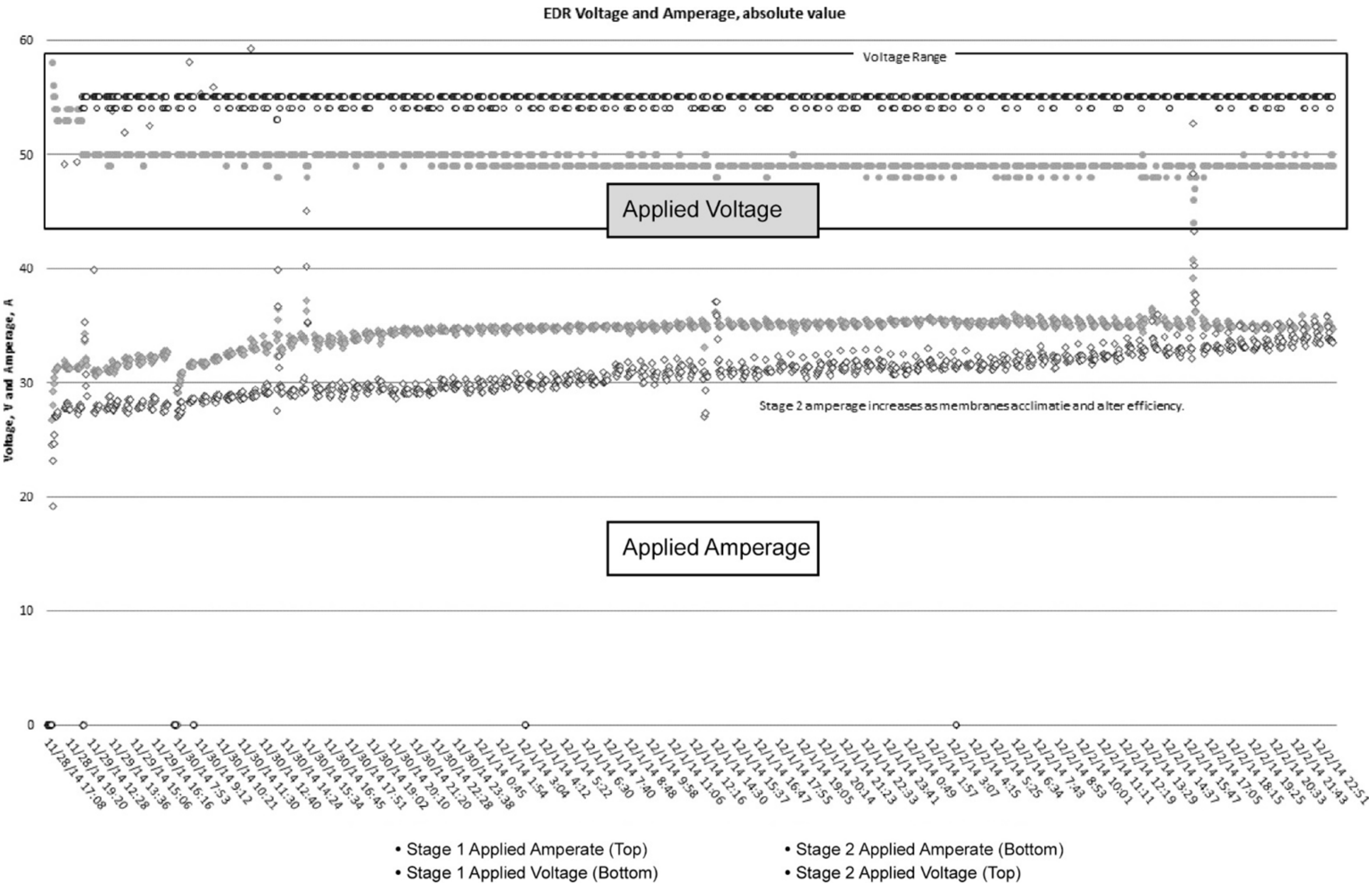


Figure 13. EDR voltage and amperage during Phase 1 operation.

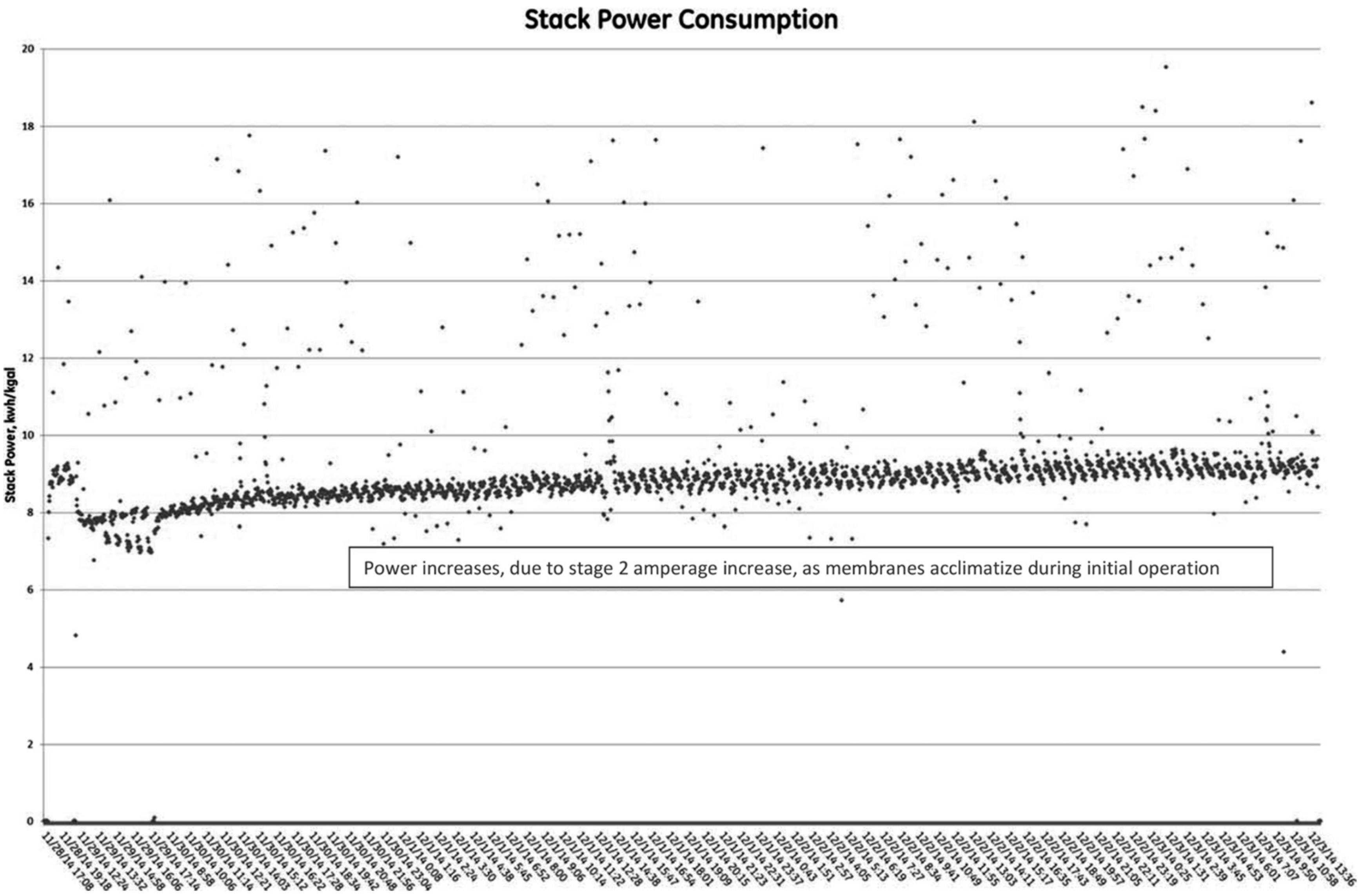


Figure 14. EDR stack power consumption during Phase 1 operation.

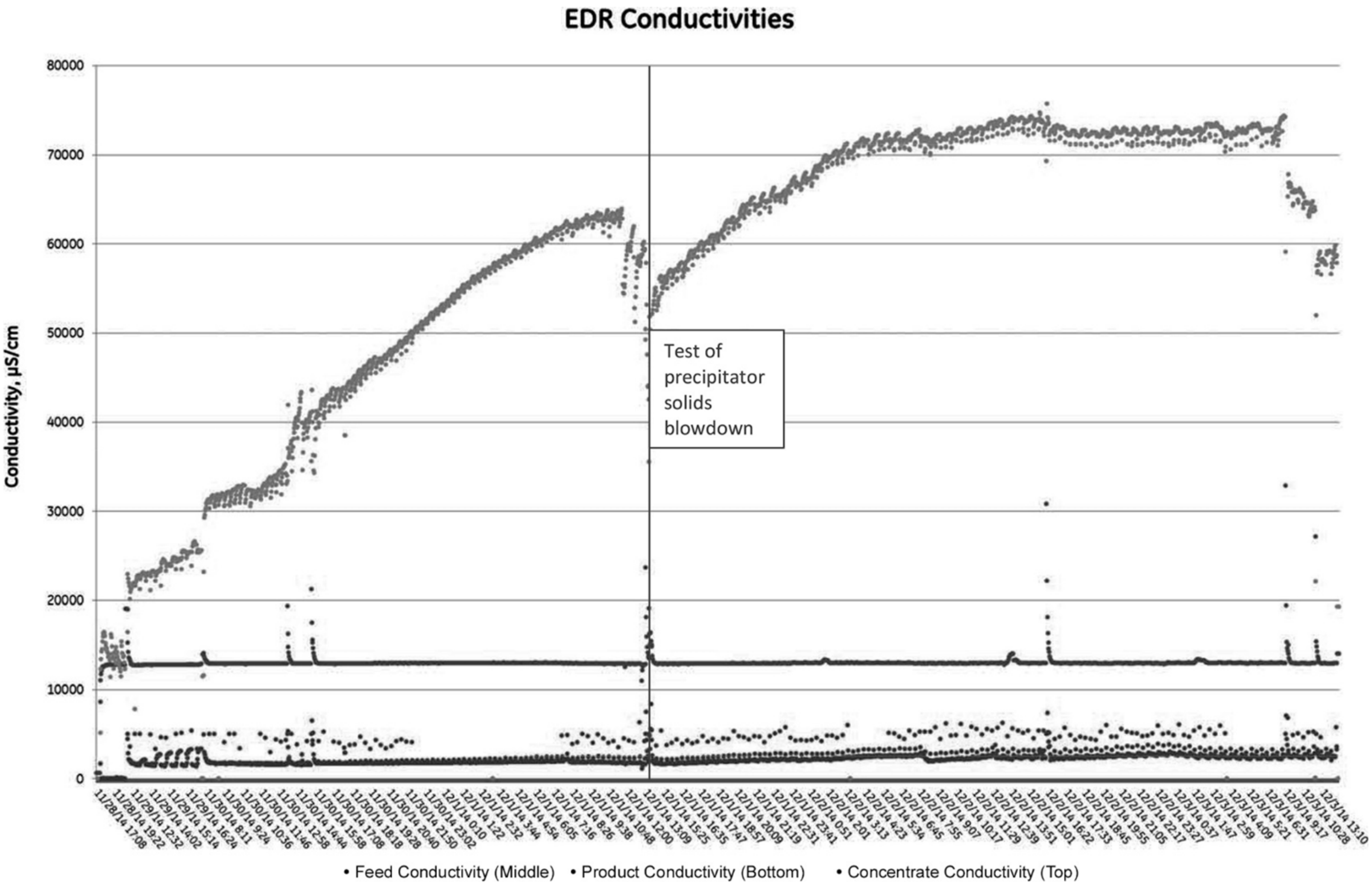


Figure 15. EDR conductivities during Phase 1 operation.

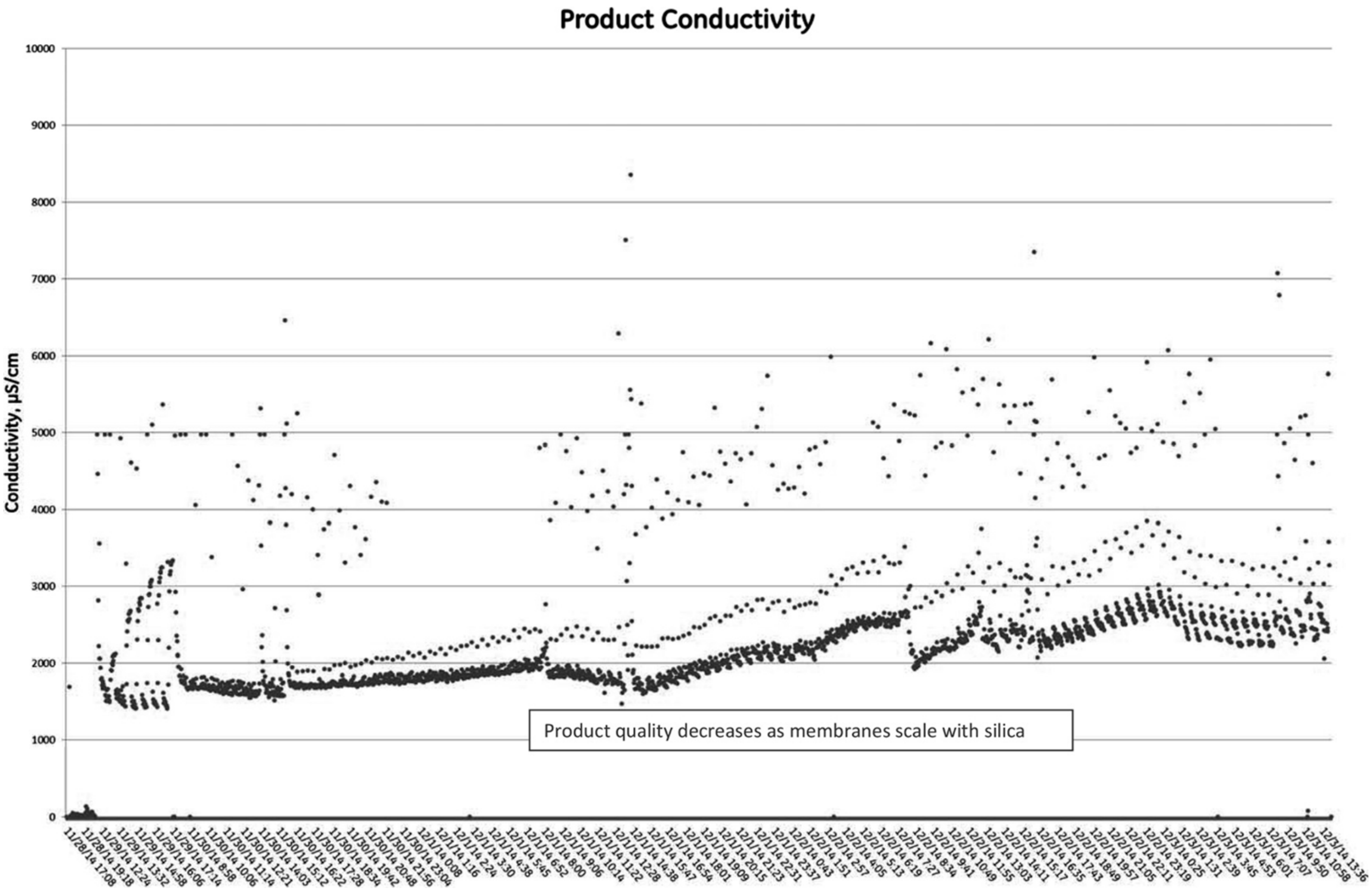
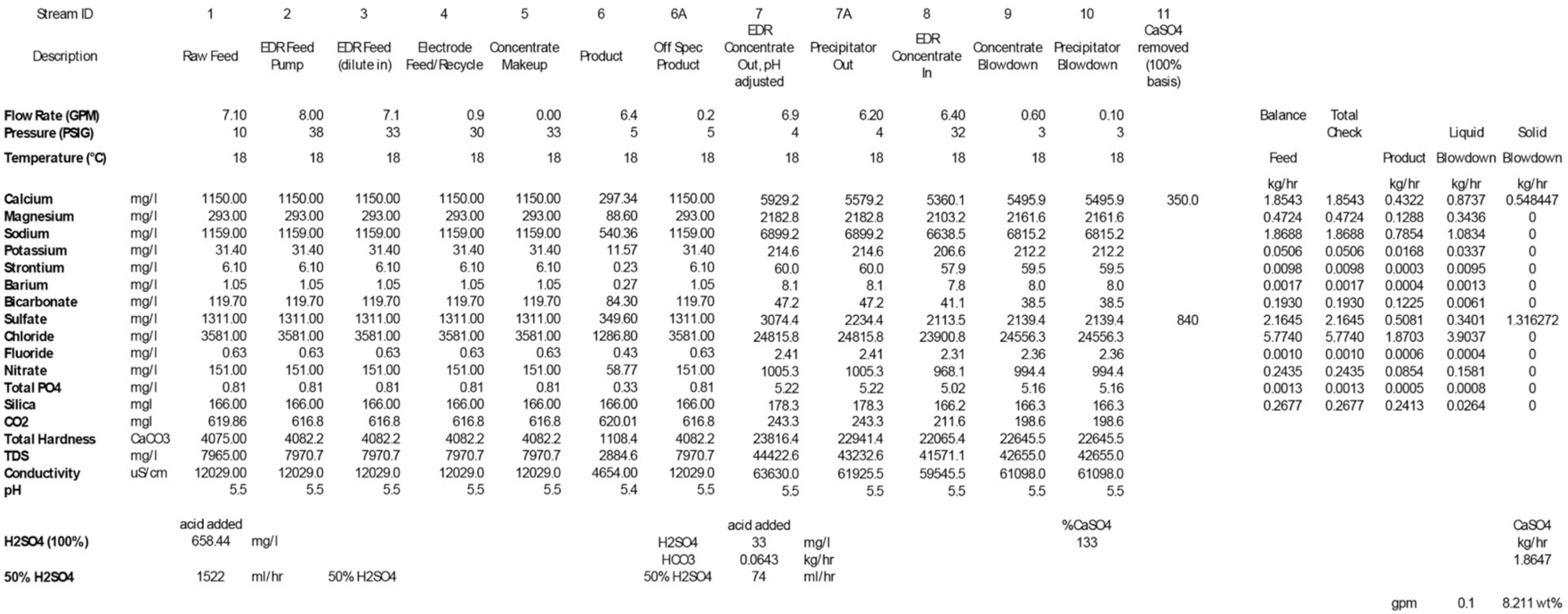
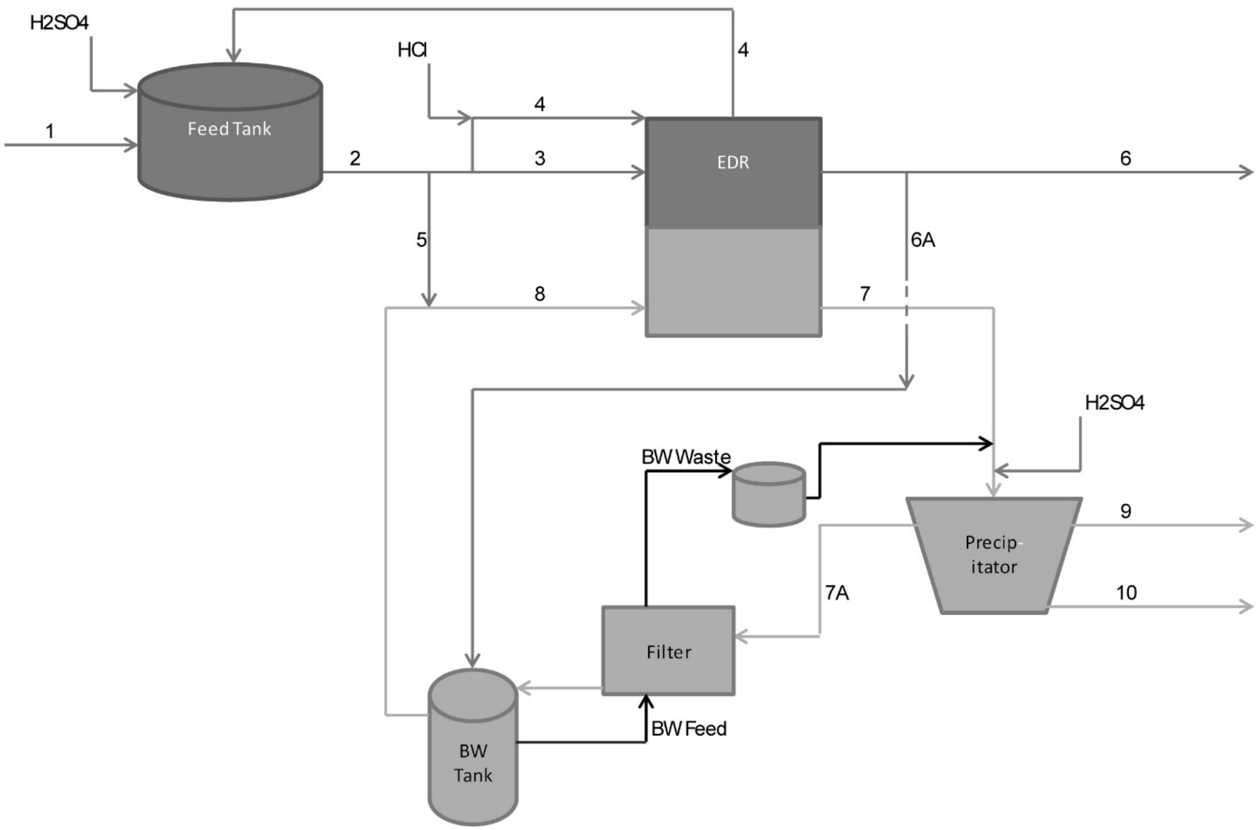


Figure 16. Phase 1 EDR product conductivity blowup.



Stream ID		1	2	3	4	5	6	6A	7	7A	8	9	10	11					
Description		Raw Feed	EDR Feed Pump	EDR Feed (dilute in)	Electrode Feed/Recycle	Concentrate Makeup	Product	Off Spec Product	EDR Concentrate Out, pH adjusted	Precipitator Out	EDR Concentrate In	Concentrate Blowdown	Precipitator Blowdown	CaSO4 removed (100% basis)					
Flow Rate (GPM)		7.10	8.00	7.1	0.9	0.00	6.4	0.2	6.9	6.20	6.40	0.60	0.10		Balance	Total			
Pressure (PSIG)		10	38	33	30	33	5	5	4	4	32	3	3			Check		Liquid	Solid
Temperature (°C)		18	18	18	18	18	18	18	18	18	18	18	18		Feed		Product	Blowdown	Blowdown
															kg/hr		kg/hr	kg/hr	kg/hr
Calcium	mg/l	1150.00	1150.00	1150.00	1150.00	1150.00	297.34	1150.00	5929.2	5579.2	5360.1	5495.9	5495.9	350.0	1.8543	1.8543	0.4322	0.8737	0.548447
Magnesium	mg/l	293.00	293.00	293.00	293.00	293.00	88.60	293.00	2182.8	2182.8	2103.2	2161.6	2161.6		0.4724	0.4724	0.1288	0.3436	0
Sodium	mg/l	1159.00	1159.00	1159.00	1159.00	1159.00	540.36	1159.00	6899.2	6899.2	6638.5	6815.2	6815.2		1.8688	1.8688	0.7854	1.0834	0
Potassium	mg/l	31.40	31.40	31.40	31.40	31.40	11.57	31.40	214.6	214.6	206.6	212.2	212.2		0.0506	0.0506	0.0168	0.0337	0
Strontium	mg/l	6.10	6.10	6.10	6.10	6.10	0.23	6.10	60.0	60.0	57.9	59.5	59.5		0.0098	0.0098	0.0003	0.0095	0
Barium	mg/l	1.05	1.05	1.05	1.05	1.05	0.27	1.05	8.1	8.1	7.8	8.0	8.0		0.0017	0.0017	0.0004	0.0013	0
Bicarbonate	mg/l	119.70	119.70	119.70	119.70	119.70	84.30	119.70	47.2	47.2	41.1	38.5	38.5		0.1930	0.1930	0.1225	0.0061	0
Sulfate	mg/l	1311.00	1311.00	1311.00	1311.00	1311.00	349.60	1311.00	3074.4	2234.4	2113.5	2139.4	2139.4	840	2.1645	2.1645	0.5081	0.3401	1.316272
Chloride	mg/l	3581.00	3581.00	3581.00	3581.00	3581.00	1286.80	3581.00	24815.8	24815.8	23900.8	24556.3	24556.3		5.7740	5.7740	1.8703	3.9037	0
Fluoride	mg/l	0.63	0.63	0.63	0.63	0.63	0.43	0.63	2.41	2.41	2.31	2.36	2.36		0.0010	0.0010	0.0006	0.0004	0
Nitrate	mg/l	151.00	151.00	151.00	151.00	151.00	58.77	151.00	1005.3	1005.3	968.1	994.4	994.4		0.2435	0.2435	0.0854	0.1581	0
Total PO4	mg/l	0.81	0.81	0.81	0.81	0.81	0.33	0.81	5.22	5.22	5.02	5.16	5.16		0.0013	0.0013	0.0005	0.0008	0
Silica	mgl	166.00	166.00	166.00	166.00	166.00	166.00	166.00	178.3	178.3	166.2	166.3	166.3		0.2677	0.2677	0.2413	0.0264	0
CO2	mgl	619.86	616.8	616.8	616.8	616.8	620.01	616.8	243.3	243.3	211.6	198.6	198.6						
Total Hardness	CaCO3	4075.00	4082.2	4082.2	4082.2	4082.2	1108.4	4082.2	23816.4	22941.4	22065.4	22645.5	22645.5						
TDS	mg/l	7965.00	7970.7	7970.7	7970.7	7970.7	2884.6	7970.7	44422.6	43232.6	41571.1	42655.0	42655.0						
Conductivity	uS/cm	12029.00	12029.0	12029.0	12029.0	12029.0	4654.00	12029.0	63630.0	61925.5	59545.5	61098.0	61098.0						
pH		5.5	5.5	5.5	5.5	5.5	5.4	5.5	5.5	5.5	5.5	5.5	5.5						
		acid added							acid added				%CaSO4						CaSO4
H2SO4 (100%)		658.44	mg/l					H2SO4	33	mg/l			133						kg/hr
								HCO3	0.0643	kg/hr									1.8647
50% H2SO4		1522	ml/hr	50% H2SO4				50% H2SO4	74	ml/hr									
																	gpm	0.1	8.211 wt%

Figure 17. Mass balance for Phase 1 operations.



Stream ID	Description	Flow Rate (gpm)
1	Raw Feed	7.1
2	EDR Feed	8.0
3	Dilute In	7.1
4	Electrode Feed/Recycle	0.9
5	Concentrate Makeup	0.0
6	Product	6.4
6A	Off Spec Product	0.2
7	Concentrate Out	6.9
7A	Precipitator Effluent	6.2
8	Concentrate In	6.4
9	Concentrate Blowdown	0.6
10	Precipitator Blowdown	0.1

Figure 18. Simplified process flow diagram for Phase 1 operations.

4.1.5 Salt Rejection

Figure 19 shows the EDR salt rejection during Phase 1 testing. The values were calculated based on the in-line conductivity probe using the TDS/conductivity empirical factors described in Section 3.4.1. As shown in Figure 19, there is considerable variability in the data, which is to be expected considering the number of operational upsets encountered during this phase. However, the average salt rejection throughout Phase 1 was 67 percent.

4.2 Phase 2: Process Stabilization Discussion

Phase 2 represents testing from January 28 to May 6, 2015. Once the silica problem was solved, the team was able to review other issues within the rest of the process. In order to operate at steady state, the next order of business was to balance flows within the AquaSel pilot. During this step, the Menifee Desalter operated at 70 percent recovery with the AquaSel set to 90 percent recovery.

Because the precipitator’s outer ring functions essentially as an upflow clarifier, a consistent, slow flow velocity is crucial. Flow upsets could cause high levels of solids to float to the top of the precipitator and get recycled back into the EDR stack.

While the EDR concentrate flow is essentially constant, there are two intermittent flows that impact precipitator performance: off-spec product and candle filter backwash. Due to membrane fouling, off-spec product flows increased, causing imbalances in the system. Furthermore, the candle filter backwash flow was being recycled at too high a rate to the precipitator, causing an additional temporary influx of water to the precipitator. The effect of these varying recycle flows was periodic hydraulic overloading of the precipitator, further resulting in unintended loss of solids from the high carryover rate. The loss of solids from the precipitator overloaded the candle filters. Figure 20 shows the candle filters and the accumulation of solids on the upper ring of the precipitator.

To correct the issue of intermittent flows, the GE process team altered flows by repurposing a tank to collect and distribute the intermittent flows to the precipitator more slowly and over a longer period of time. Additionally, in mid-March, the pretreatment acid addition was stopped. With these process changes, the operation stabilized. Shutdowns became less frequent, and the system was able to run continuously for a few days without initiating any alarms.

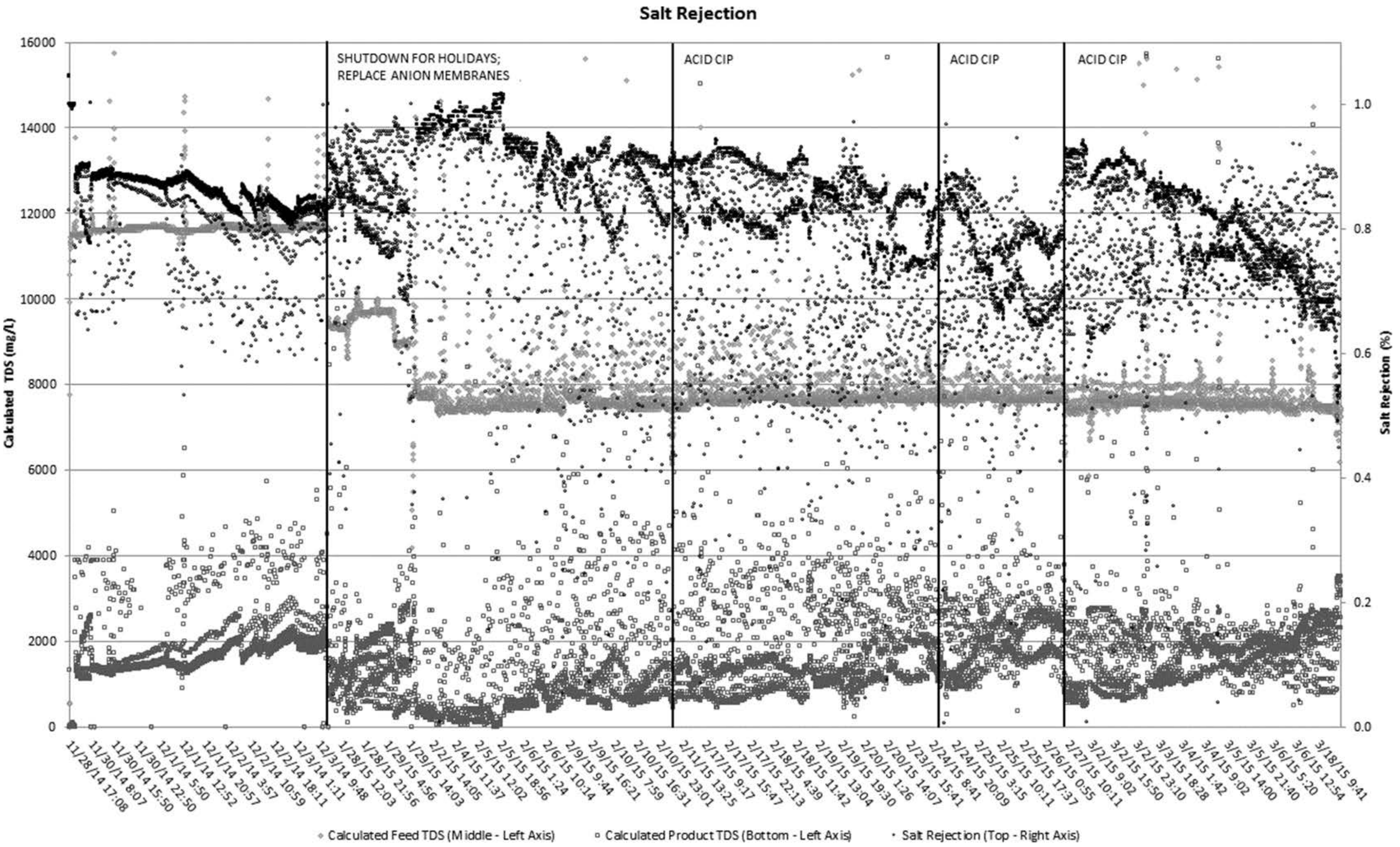


Figure 19. Salt rejection during Phase 1 operations.

Figure 20. Calcium sulfate deposition in the precipitator and the candle filter.

4.2.1 New Membrane Installation – April 28, 2015

Once reliable multi-day operation was achieved, a decision was made to replace the anion membranes in the EDR stack. Now that the process was stable, it was important to show how a new unit would be expected to operate. The goal of replacing these membranes was to improve product water quality. Since the previous membranes had been fouled by the silica, the product water was not meeting expectations and could not have been expected to.

4.2.2 Testing Objectives

Once the silica fouling issue was resolved, it was critical to stabilize flows to prevent unintentional solids overflows from the precipitator. During this phase of the project, very few grab samples were collected as the purpose of Phase 2 was stabilization. Therefore, the objectives for Phase 2 were as follows:

- Mitigate surge flows into the precipitator feed.
- Prevent solids overflow from the precipitator.
- Optimize precipitator effluent candle and cartridge filter operation.

4.2.3 Feed Water Quality

Table 7 shows the Menifee Desalter concentrate water quality during Phase 2.

Table 7. Menifee Desalter Concentrate – EDR Feedwater Quality[1]

Parameter	Units	Value
Alkalinity as $CaCO_3$	mg/L	3,312
Total Dissolved Solids (TDS)	mg/L	5,623
Sulfate	mg/L	422
Sodium	mg/L	627
Calcium	mg/L	976
Magnesium	mg/L	212
Chloride	mg/L	2,350
Total Silica	mg/L	148

[1] Menifee Desalter was operated at a recovery of 70 percent.

4.2.4 Hydraulic Performance

As described in Section 4.2, the process layout was modified to manage surge flows. Furthermore, acid pretreatment addition was discontinued at the beginning of Phase 2. As Figure 21 illustrates, the system pressures during Phase 2 quickly stabilized and remained fairly consistent from mid-March through April. When the EDR anion membranes were replaced on April 29th, the pressures initially dropped but then resumed the previous trend and increased to approximately 35 psi.

4.2.5 Electric Performance

A distinct amperage drop occurred when Phase 2 was initiated and pretreatment stopped in mid-March. Some variability remained throughout Phase 2, but at a much lower amperage than in Phase 1. Similarly, Stage 1 and Stage 2 voltages dropped by approximately 10 volts and remained fairly constant throughout Phase 2. Figure 22 shows a similar decrease in stack power consumption from 7 kilowatt-hours per thousand gallons (kWh/kgal) in Phase 1 to about 4 kWh/kgal in Phase 2.

4.2.6 Water Quality

As stated above, very few grab samples were collected during this phase. However, one sample was collected on April 17, 2015, to gauge the process conditions at that point. The results from the GE Woodlands analytical laboratory are listed in Table 8.

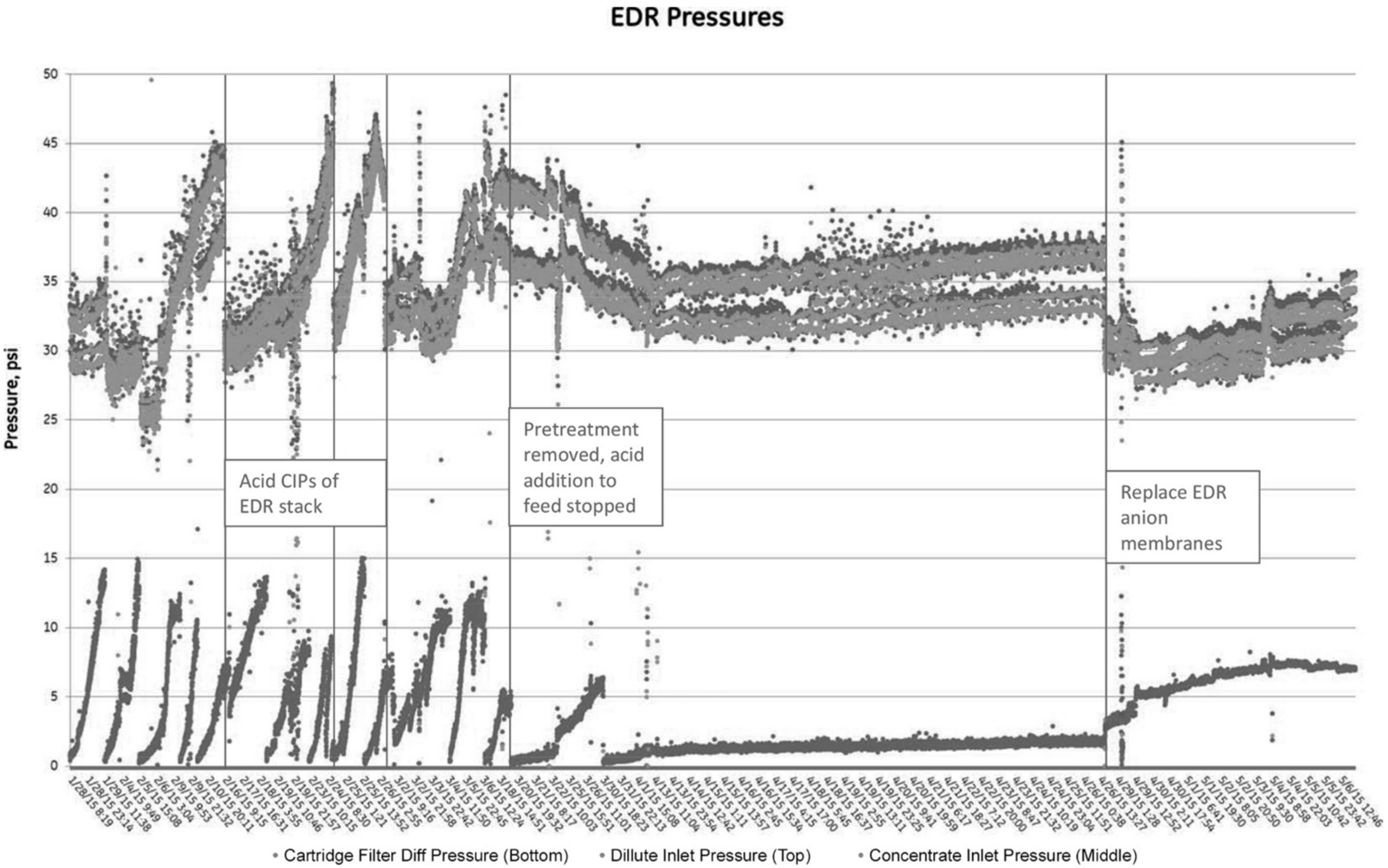


Figure 21. EDR pressures during Phase 2 operations.

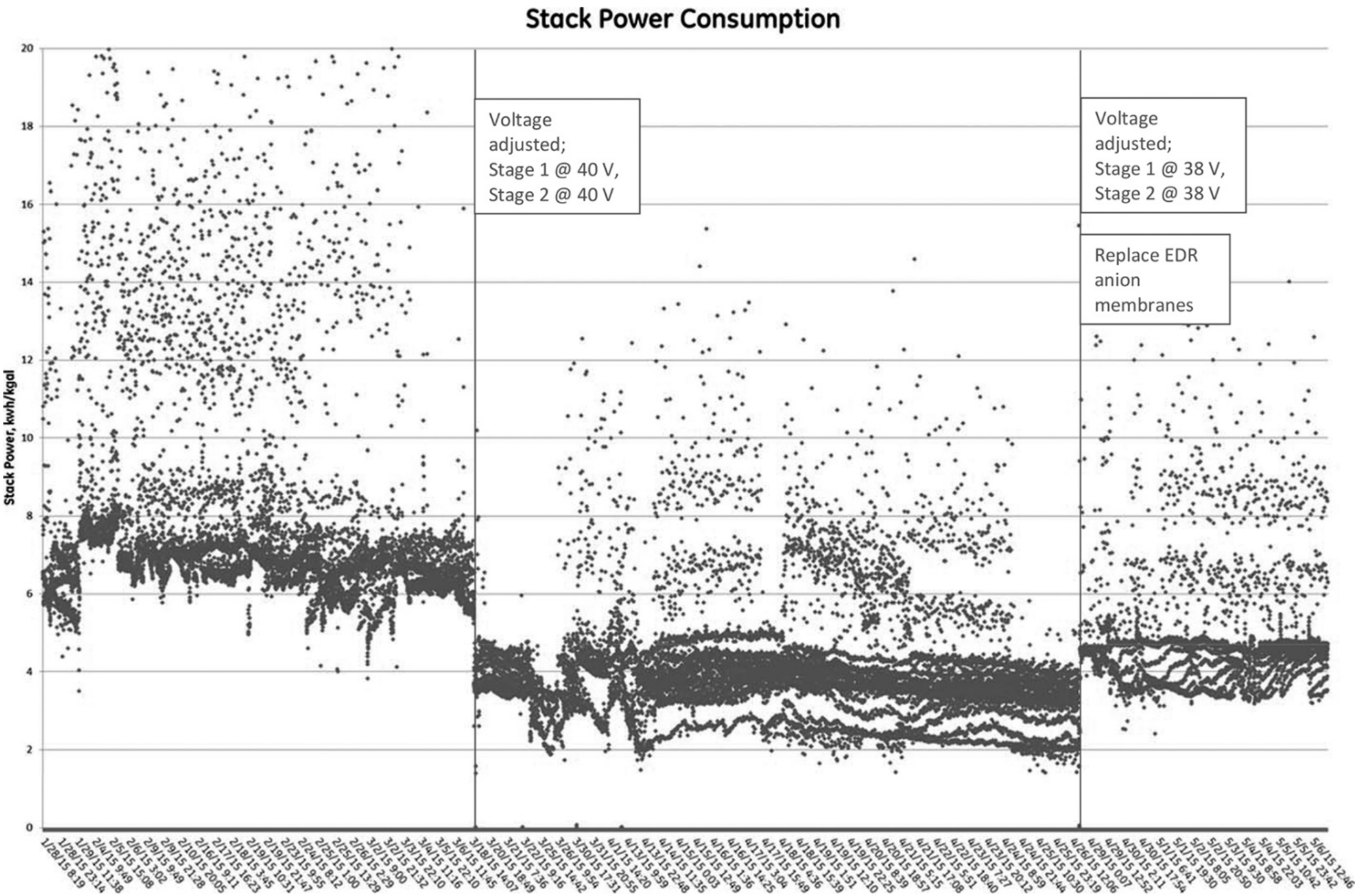


Figure 22. EDR stack power consumption during Phase 2 operation.

Table 8. Process Stream Water Quality Measured During Phase 2

Parameter	Units	Raw	Product	EDR Conc.	Precip. Effluent
pH	units	7.7	7.5	6.6	6.5
Conductivity	µS/cm	8,760	4,130	56,600	54,500
Calcium	mg/L	976	125	8,040	7,360
Magnesium	mg/L	212	40.5	2,044	1,913
Sodium	mg/L	627	435	2,530	2,400
Potassium	mg/L	21.5	10.9	125	118
Sulfate	mg/L	422	47.9	1,750	1,570
Chloride	mg/L	2,350	949	21,500	22,300
Nitrate	mg/L	119	29.1	1,100	1,100
Total silica	mg/L	148	149	107	107
Alkalinity as $CaCO_3$	mg/L	3,320	480	28,600	26,300
TDS	mg/L	5,618	2,256	38,120	37,602

As shown in Table 8, Phase 2 did not meet the product conductivity goal of less than 3,200 µS/cm. Similarly, the calcium and sulfate removal was not as expected compared to Phase 1. Figures 23 and 24 show plots of the conductivity measured during Phase 2.

Additionally, GE, using their WATSYS system, prepared a mass balance analysis for operations during Phase 2 (Figure 25). Figure 26 shows the simplified process flow diagram showing the Phase 2 operation schematic.

4.2.7 Salt Rejection

At the beginning of Phase 2, calculated product TDS increased while calculated feed TDS remained fairly constant, indicating an obvious decrease in salt rejection. The calculated product TDS continued to climb until the anion membranes were replaced on April 29th. The resulting average salt rejection during Phase 2 was 41 percent with a maximum of around 80 percent and a minimum of less than 10 percent. However, once the anion membranes were replaced, the average salt rejection went up to 57 percent.

4.3 Phase 3 Results

Phase 3 represents testing from June 19 to July 31, 2015, during which the operating parameters were held as stable as possible. The objective of this test period was to obtain a set of data for a single set of operating conditions from the unit, from which performance estimates could be made.

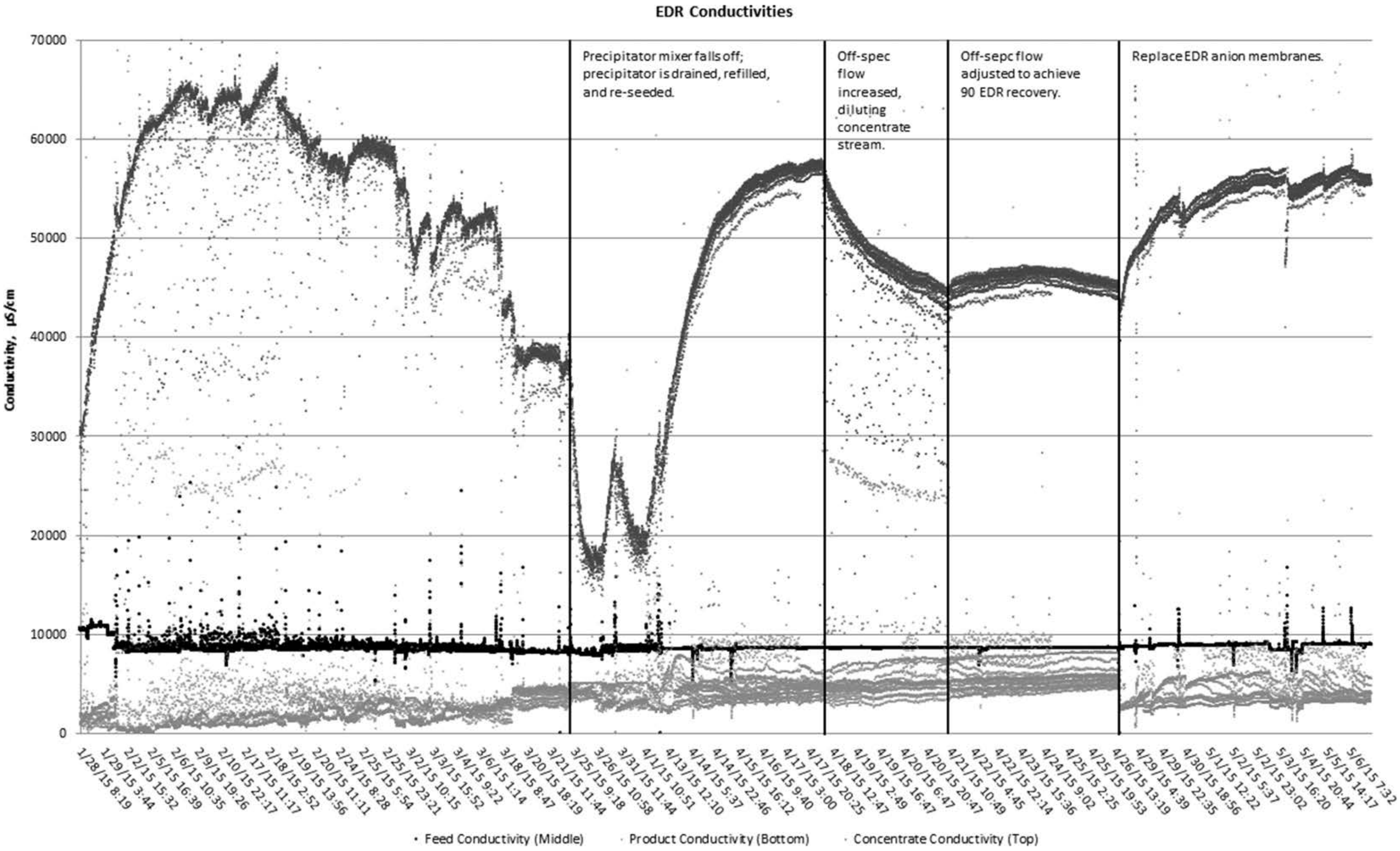


Figure 23. EDR conductivities during Phase 2 operation.

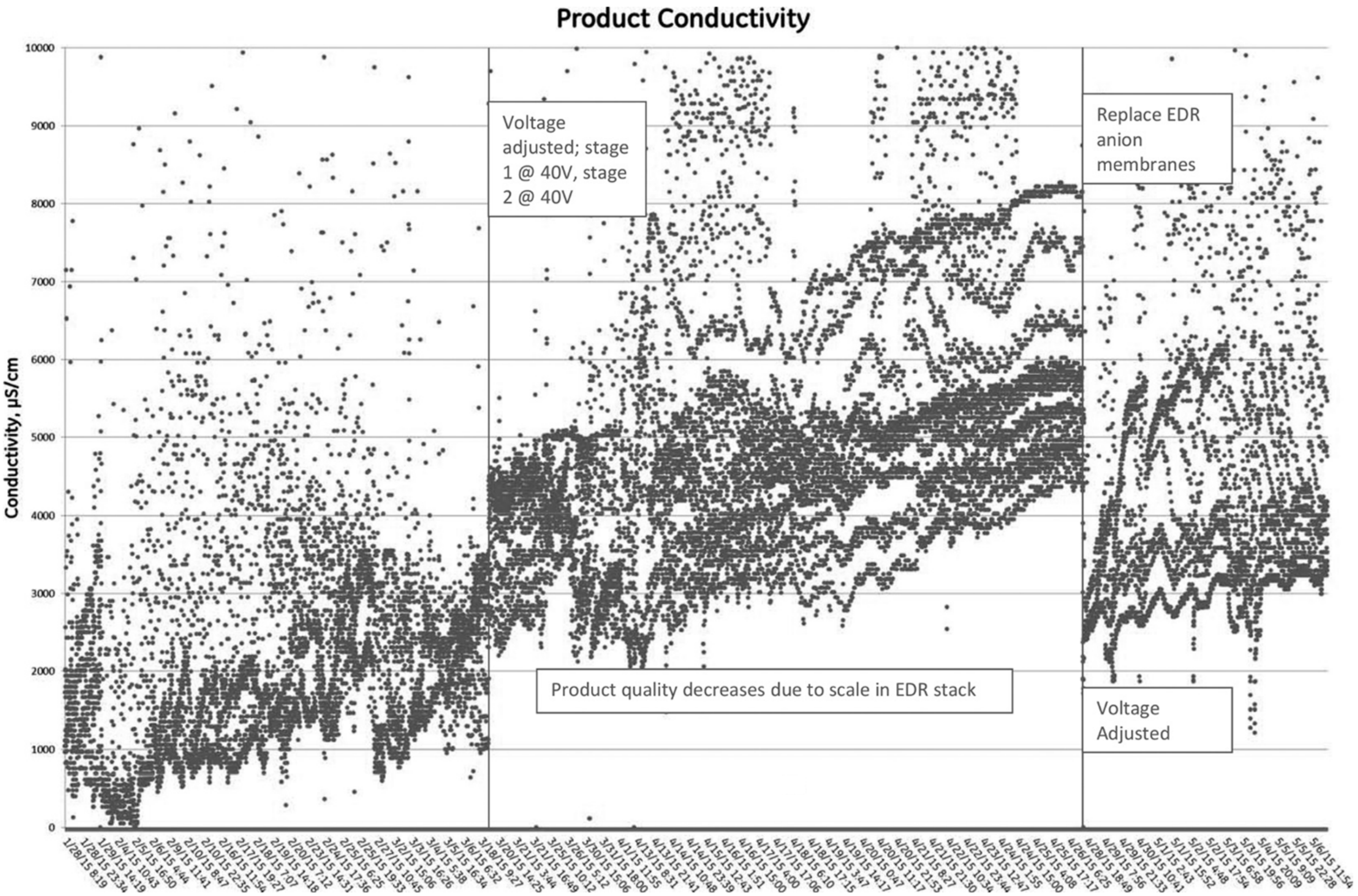
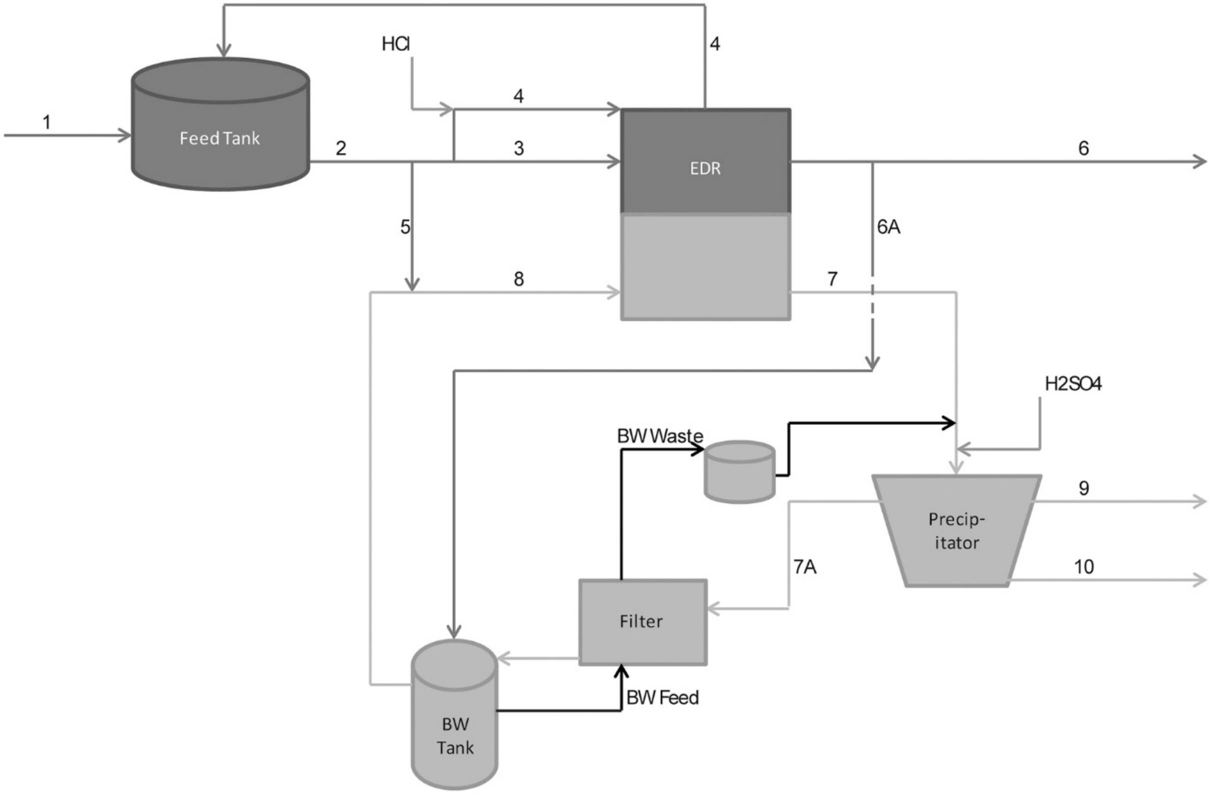


Figure 24. Phase 2 EDR product conductivity blowup.

Stream ID		1	2	3	4	5	6	6A	7	7A	8	9	10	11	Balance Feed kg/hr	Total Check	Product kg/hr	Liquid Blowdown kg/hr	Solid Blowdown kg/hr
Description		Raw Feed	EDR Feed Pump	EDR Feed (dilute in)	Electrode Feed/Recycle	Concentrate Makeup	Product	Off Spec Product	EDR Concentrate Out, pH adjusted	Precipitator Out	EDR Concentrate In	Concentrate Blowdown	Precipitator Blowdown	CaSO4 removed (100% basis)					
Flow Rate (GPM)		7.10	8.00	7.1	0.9	0.00	6.4	0.2	6.7	6.00	6.20	0.60	0.10						
Pressure (PSIG)		10	38	33	30	33	5	5	4	4	32	3	3						
Temperature (°C)		18	18	18	18	18	18	18	18	18	18	18	18						
Calcium	mg/l	976.00	976.00	976.00	976.00	976.00	124.80	976.00	7251.1	7066.1	6799.8	6993.9	6993.9	185.0	1.5737	1.5747	0.1814	1.1118	0.28149
Magnesium	mg/l	211.70	211.70	211.70	211.70	211.70	40.50	211.70	1792.5	1792.5	1726.2	1776.7	1776.7		0.3413	0.3413	0.0589	0.2824	0
Sodium	mg/l	627.00	627.00	627.00	627.00	627.00	435.00	627.00	2429.3	2429.3	2325.9	2382.5	2382.5		1.0110	1.0110	0.6322	0.3788	0
Potassium	mg/l	21.50	21.50	21.50	21.50	21.50	10.90	21.50	120.2	120.2	115.5	118.6	118.6		0.0347	0.0347	0.0158	0.0189	0
Strontium	mg/l	6.40	6.40	6.40	6.40	6.40	0.80	6.40	58.0	58.0	55.9	57.5	57.5		0.0103	0.0103	0.0012	0.0091	0
Barium	mg/l	0.73	0.73	0.73	0.73	0.73	0.07	0.73	7.0	7.0	6.8	7.0	7.0		0.0012	0.0012	0.0001	0.0011	0
Bicarbonate	mg/l	735.70	735.70	735.70	735.70	735.70	466.00	735.70	814.9	814.9	759.2	760.0	760.0		1.1863	1.1863	0.6773	0.1208	0
Sulfate	mg/l	422.00	422.00	422.00	422.00	422.00	47.90	422.00	1990.2	1546.2	1479.5	1514.8	1514.8	444	0.9860	0.9860	0.0696	0.2408	0.675577
Chloride	mg/l	2350.00	2350.00	2350.00	2350.00	2350.00	949.00	2350.00	15334.7	15334.7	14746.2	15159.4	15159.4		3.7892	3.7892	1.3793	2.4099	0
Fluoride	mg/l	0.75	0.75	0.75	0.75	0.75	0.60	0.75	2.39	2.39	2.31	2.36	2.36		0.0012	0.0012	0.0009	0.0004	0
Nitrate	mg/l	119.00	119.00	119.00	119.00	119.00	29.10	119.00	950.0	950.0	914.6	941.1	941.1		0.1919	0.1919	0.0423	0.1496	0
Total PO4	mg/l	4.10	4.10	4.10	4.10	4.10	2.50	4.10	18.83	18.83	18.04	18.50	18.50		0.0066	0.0066	0.0036	0.0029	0
Silica	mgl	148.00	148.00	148.00	148.00	148.00	148.00	148.00	158.8	158.8	147.8	147.8	147.8		0.2386	0.2386	0.2151	0.0235	0
CO2	mgl	22.03	22.0	22.0	22.0	22.0	22.29	22.0	320.1	404.5	376.9	377.3	377.3						
Total Hardness	CaCO3	3312.20	3312.2	3312.2	3312.2	3312.2	478.9	3312.2	25512.8	25050.3	24111.3	24804.6	24804.6						
TDS	mg/l	5622.9	5622.9	5622.9	5622.9	5622.9	2255.2	5622.9	30927.9	30298.9	29097.6	29880.1	29880.1						
Conductivity	uS/cm	8760.00	8760.0	8760.0	8760.0	8760.0	4130.00	8760.0	63630.0	62335.9	59864.4	61474.2	61474.2						
pH		7.7	7.7	7.7	7.7	7.7	7.5	7.7	6.6	6.5	6.5	6.5	6.5						
									acid added				%CaSO4						CaSO4
H2SO4 (100%)								H2SO4	205	mg/l			138.7						kg/hr
								HCO3	0.3882	kg/hr									0.9571
50% H2SO4								50% H2SO4	447	ml/hr									
																	gpm	0.1	4.2143 wt%

Figure 25. Mass balance for Phase 2 operations.



Stream ID	Description	Flow Rate (gpm)
1	Raw Feed	7.1
2	EDR Feed	8.0
3	Dilute In	7.1
4	Electrode Feed/Recycle	0.9
5	Concentrate Makeup	0.0
6	Product	6.4
6A	Off Spec Product	0.2
7	Concentrate Out	6.7
7A	Precipitator Effluent	6.0
8	Concentrate In	6.2
9	Concentrate Blowdown	0.6
10	Precipitator Blowdown	0.1

Figure 26. Simplified process flow diagram for Phase 2 operations.

4.3.1 Testing Objectives

After the system was deemed stable, the objectives of the remaining testing were as follows:

- Achieve consistent operation of the AquaSel system on Menifee Desalter concentrate.
- Maintain constant operation of the AquaSel system and demonstrate stable solids production and dissolved solids mass balance, as well as constant product water quality (conductivity).
- Determine the frequency of CIP procedures needed to maintain the performance of the EDR membrane stack.

4.3.2 Feed Water Quality

Table 9 summarizes the average and maximum values for individual water quality parameters measured throughout the study. The table also includes the number of samples for which data were obtained. The average value presented in the table represents the average of the laboratory samples collected during the pilot study.

Table 9. Menifee Desalter Concentrate – EDR Feedwater Quality[1]

Parameter	Units	# of Samples	Average	Maximum
Alkalinity as $CaCO_3$	mg/L	16	665	700
Alkalinity as HCO_3	mg/L	16	809	860
Total dissolved solids (TDS)	mg/L	16	7,713	9,500
Total organic carbon	mg/L	16	3.4	5.7
Total suspended solids [2]	mg/L	16	6.8	11
Sulfate	mg/L	16	489	520
Sodium	mg/L	16	774	840
Calcium	mg/L	16	1,079	1,180
Magnesium	mg/L	16	261	320
Chloride	mg/L	16	2,952	3,200
Total silica	mg/L	16	129	140

[1] Menifee Desalter was operated at a recovery of 75 percent.

[2] Of the 16 samples, two were less than 3 mg/L or below detect.

The Phase 3 test ran virtually continuously for 962 operating hours from the final recovery adjustment until the end of the study. During this time, 16 sets of samples were collected and analyzed in the laboratory. Key results of these analyses were viewed in association with operating data at the times of sampling. Analytical results obtained during Phase 3 were used to develop a representative mass balance.

4.3.3 Hydraulic Performance

GE monitored the performance of the EDR remotely. A sample chart of flows and pressures for the EDR process is shown in Figure 27.

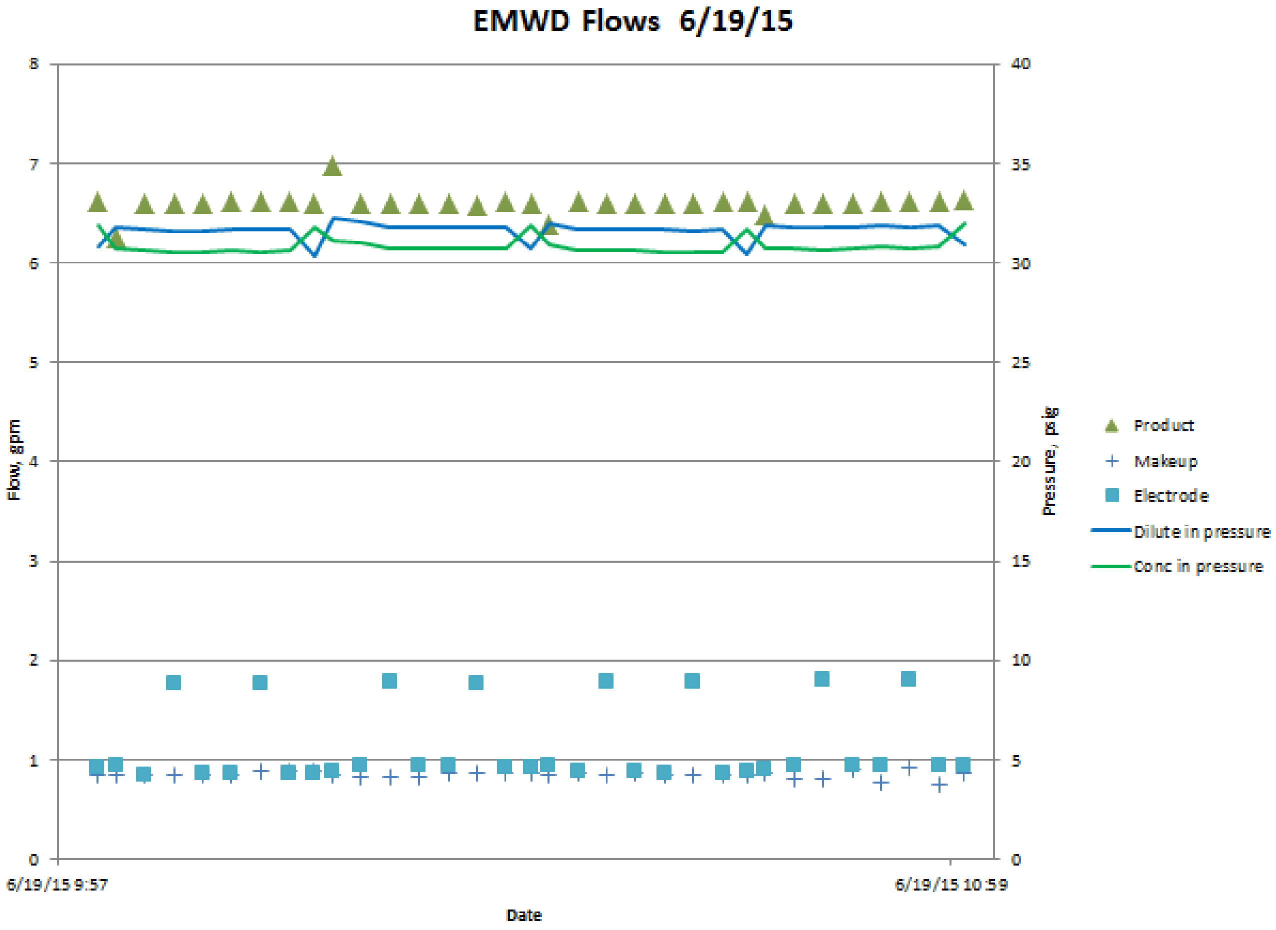


Figure 27. AquaSel flows, June 19, 2015.

This is for an operating period when the first samples were collected for analysis in Phase 3. Flows read on the left axis show the product flow rate stable at 6.6 gpm and concentrate makeup stable at 0.9 gpm. The electrode flow rate varied between 0.9 and 1.8 gpm. The figure demonstrates that the “bumping” or periodic flowing of the anode stream was working correctly.

The pressures of the dilute inlet and concentrate inlet streams to the EDR membrane stack are read on the right axis. These data show a slightly higher inlet pressure for the dilute than for the concentrate inlet, which is by design as the EDR process is intended to operate with a slightly higher inlet pressure on the diluting side of the membranes in the stack. The higher pressure on the dilute side prevents the concentrate stream from contaminating the dilute product water. Any cross-leakage between streams will flow from the higher pressure (dilute) stream to the lower pressure (concentrate) stream. The points in the process where the inlet and outlet pressures become inverted occur when the DC polarity is reversed on the membrane stack every 15 minutes. This illustrates that the control system is adjusting the inlet pressure differential as required when each polarity reversal occurs. Similar charts for flow and pressure can be found in Appendix A.2, demonstrating consistent flow and pressure during Phase 3.

4.3.4 Electric Performance

A sample chart of DC voltage and current for each of the two stages of the EDR membrane stack is shown in Figure 28.

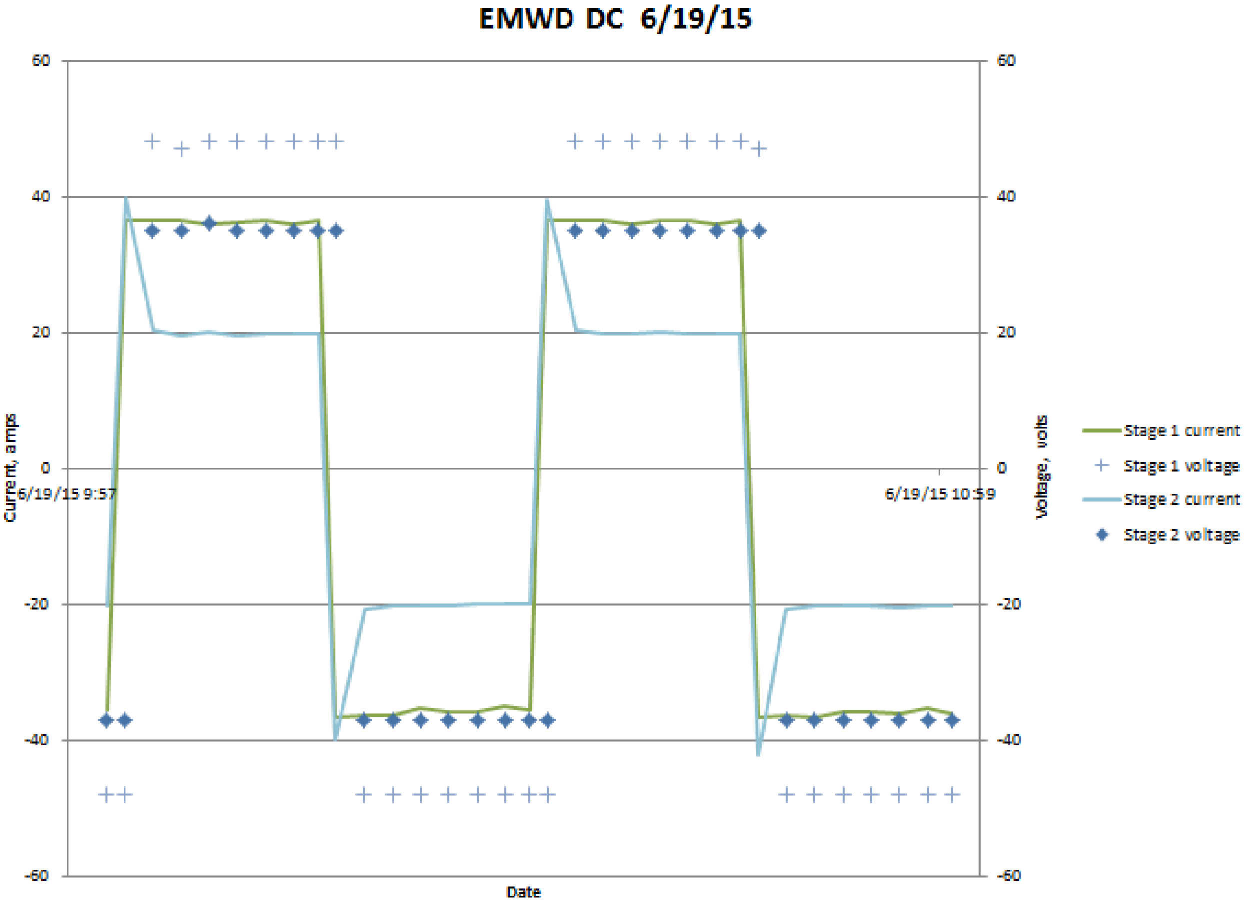


Figure 28. EDR voltage and amperage, June 19, 2015.

Individual data points are shown for DC voltages applied to electrical stage 1 and electrical stage 2. Solid lines illustrate current flow in each stage during this period of operation. The steps up and down illustrate reversal of polarity from positive to negative. During this time stage 1 voltage was stable at 48 volts and stage 2 at 36 volts on average. Stage 1 current was stable at 35 amps and stage 2 at about 20 amps. The current spikes seen in Stage 2 are from current surges at polarity reversal as the concentrating and diluting streams displace one another. This is a normal phenomenon in ER process and did not have an impact on the pilot test. Recent EDR technology upgrades utilize DC drives, which minimize current surges during reversal.

Similar charts can be found in Appendix A.3. The flatness of the current profile during each polarity as well as the product conductivity profile during each operating polarity indicates no detectable increase in electrical resistance of the membranes during Phase 3. The charts in Appendix A.3 all demonstrate that the feed water (Menifee Desalter RO brine) did not foul the EDR membranes during the final phase of testing.

4.3.5 Phase 3 Power Consumption

The greatest power consumption in the AquaSel system comes from the EDR stack and the EDR system pumping. GE's pilot system recorded the voltage and amperage across the EDR stack, which allows the power consumption to be calculated. The peripheral equipment typically draws a higher percentage of the power consumed in a pilot system than in a full-scale plant. GE provided an estimate of the power consumed based on the equipment in operation, which is presented in Table 10. Stable operation over Phase 3 resulted in a 6.2 kWh/kgal net power consumption.

Table 10. Power Consumption for AquaSel Pilot Unit

Item	Flow (gpm)	Pressure (psig)	Pump Eff (%)	Motor Eff (%)	VFD Eff (%)	kW	kWh/kgal
EDR feed pump	8.9	65	70	95	97	0.39	
EDR concentrate pump	6.4	65	70	95	97	0.28	
Filter feed pump	7	25	70	95	97	0.12	
Backwash pump	8	25	70	95		0.14	
LC pump	1	10	60	95		0.008	
Bleed pump	1	10	60	95		0.008	
Slurry pump	3	5	60	95		0.011	
HCl dosing						0.1	
H_2SO_4 dosing						0.1	
AC total						1.16	3.0
DC total						2.33	6.2

4.3.6 Water Quality

A sample chart of conductivity and calcium in each of three main process streams is shown in Figure 29. These are the feed, product, and concentrate blowdown (overflow from precipitator sent to waste).

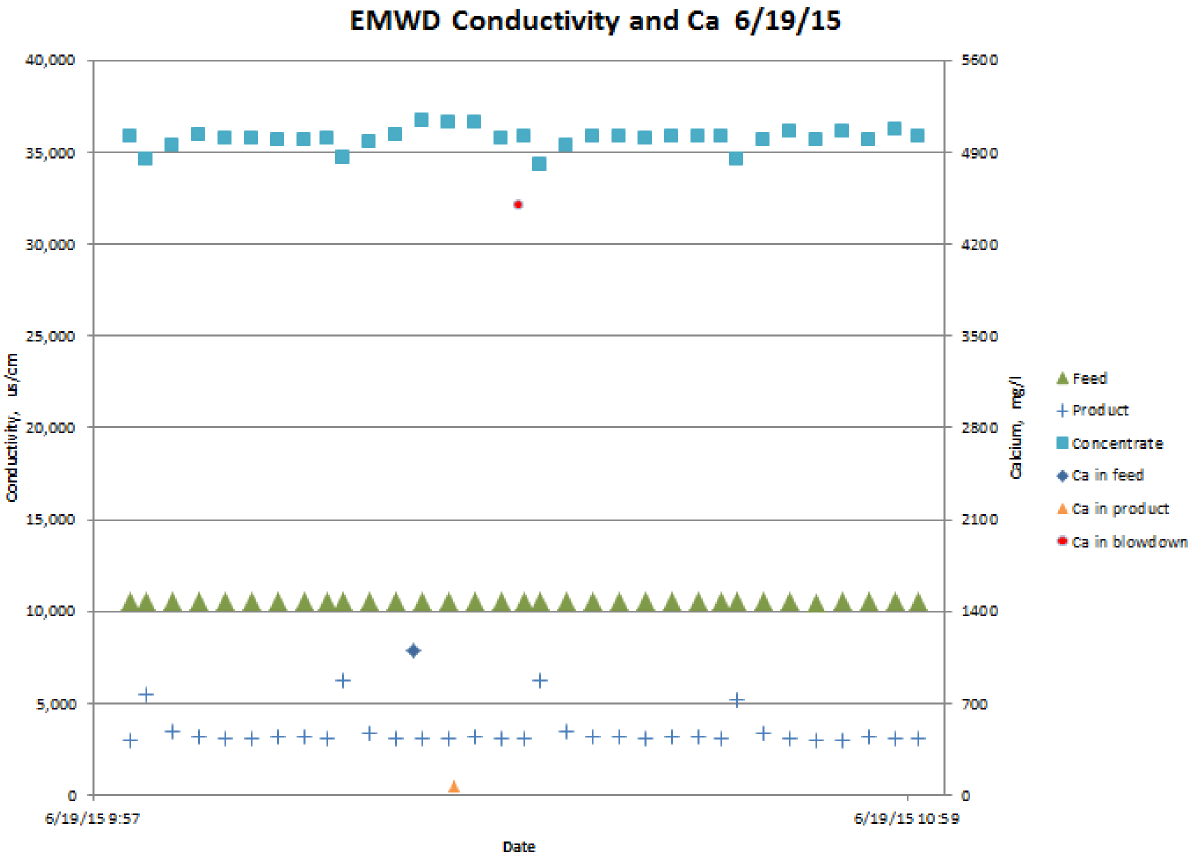


Figure 29. AquaSel conductivity and calcium, June 19, 2015.

This is for the same operating period as the previous chart. As expected, feed conductivity was stable at 10,500 µS/cm throughout (RO brine from Menifee Desalter operating at 75 percent recovery). Product water from the EDR was consistent, averaging about 3,130 µS/cm. Note that spurious readings were not included in the average. The few product conductivities >5,000 µS/cm shown in Figure 29 were associated with EDR polarity reversal, during which the positive and negative poles on the EDR membrane stack switch along with the streams in the EDR stack. Off-spec product is produced during these reversals as a normal part of EDR operation. In the pilot process, off-spec product was sent to the recirculating concentrate loop (at the inlet to precipitator) where it served as a small portion of the makeup water to the brine loop. Figure 30 shows the product conductivity averaging around 3,000 µS/cm during Phase 3. The "cloud" of data points in the 4,000 to 7,000 µS/cm range was caused by the EDR polarity reversal mentioned earlier. Figure 31 shows that the concentrate conductivity at the outlet of the EDR membrane stack was stable at around 35,000 µS/cm.

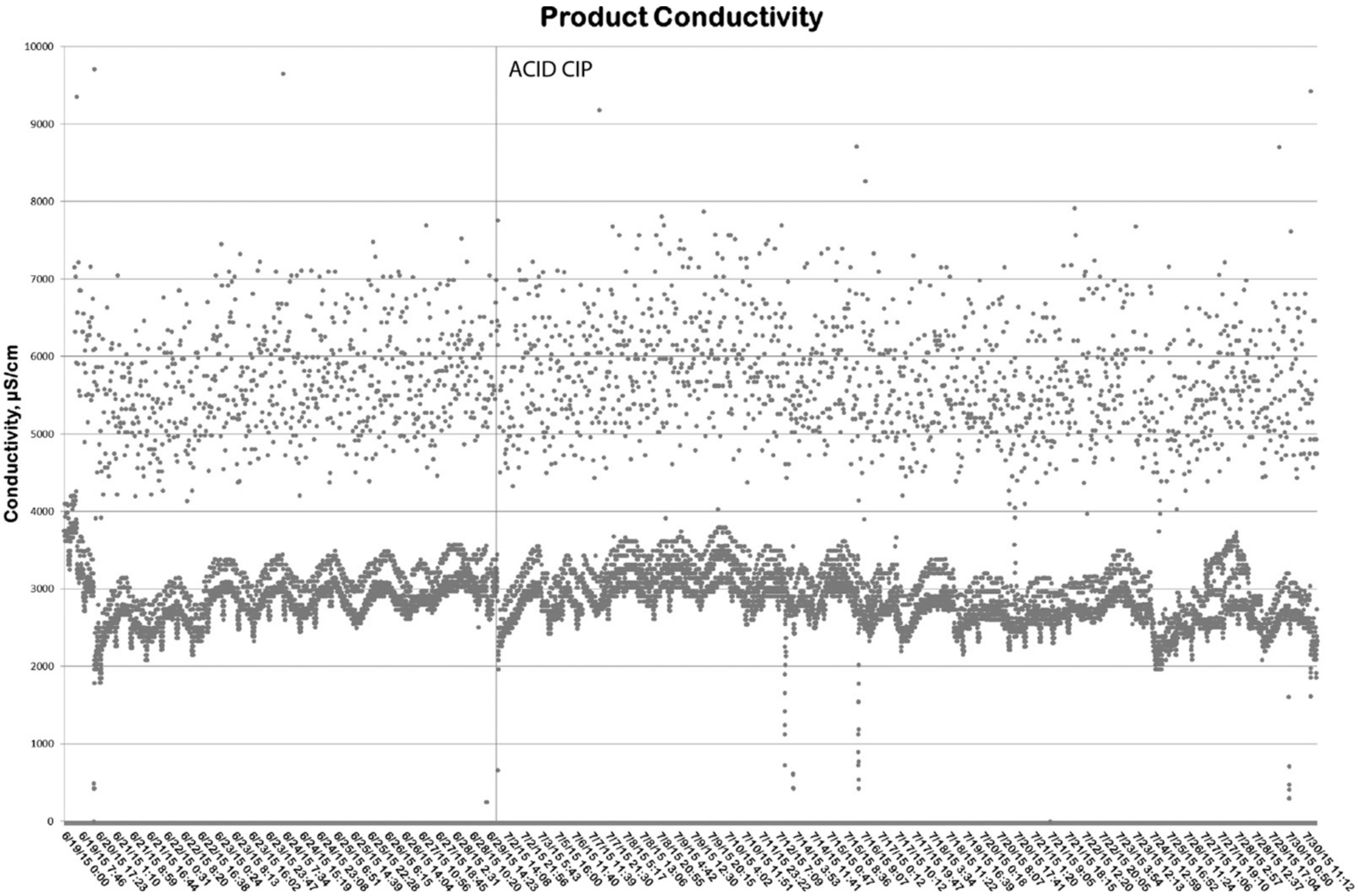


Figure 30. Phase 3 product conductivity.

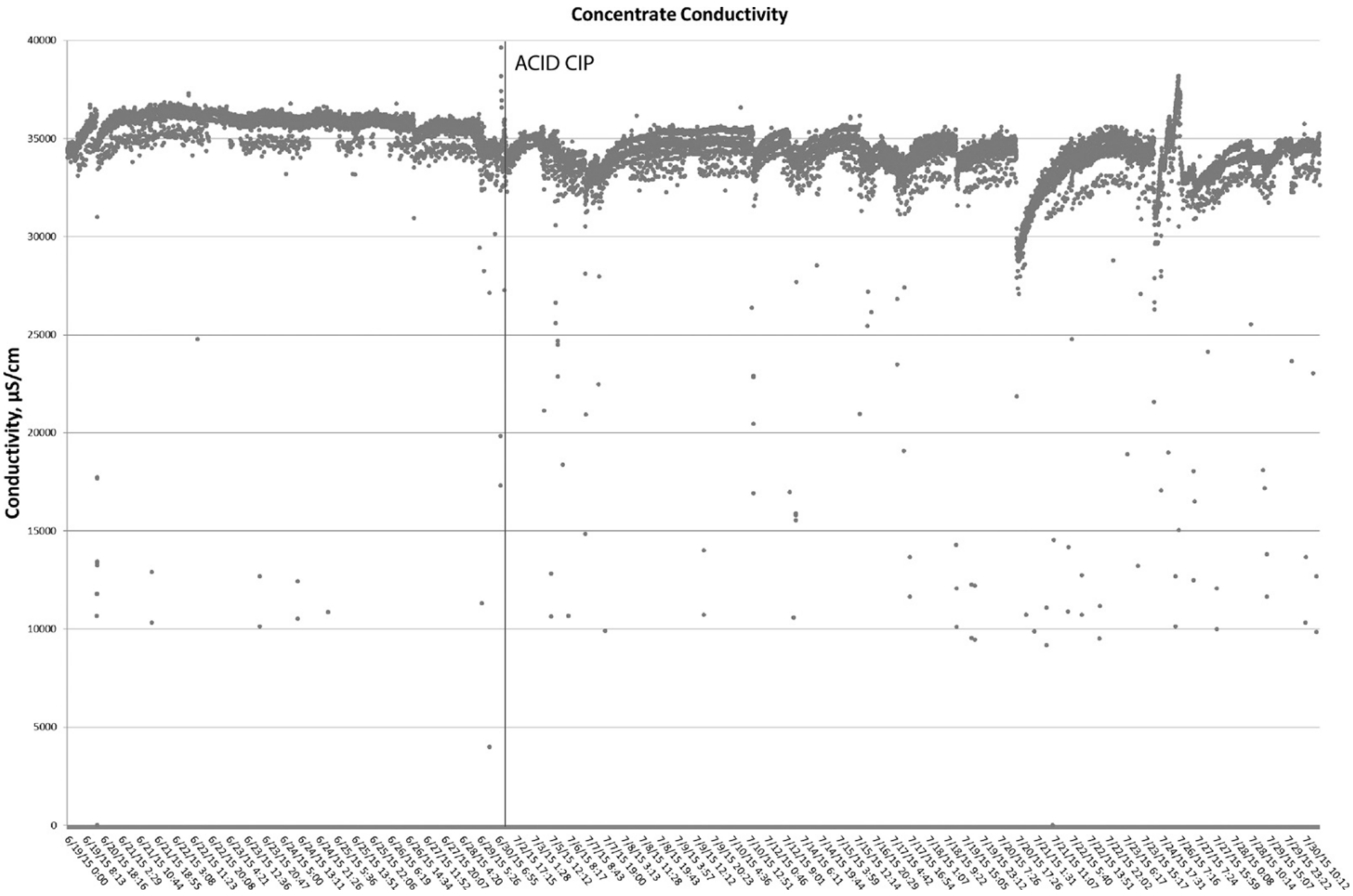


Figure 31. Phase 3 concentrate conductivity. (Note: Figure truncated to 40,000 µS/cm, although some points reach 200,000 µS/cm.)

Table 11 shows the averaged analytical results from 16 samples collected between June 19 and July 31 from the final phase of testing for Menifee Desalter feed water, AquaSel product, and precipitator effluent. Feed-water calcium was 1,079 mg/L and product calcium was 60 mg/L. Calcium in the effluent from the precipitator (after the precipitation process) was 4,177 mg/L. Similar charts for conductivity and Ca can be found in Appendix A.1, demonstrating consistent salt removal and concentration of dissolved salts in EDR effluent.

Table 11. Averaged Analytical Results – Phase 3 Performance [1]

Parameter	Units	Feed Water	AquaSel Product	AquaSel Waste
Alkalinity as $CaCO_3$	mg/L	665	336	1,093
Alkalinity as HCO_3	mg/L	809	394	
Total dissolved solids (TDS)	mg/L	7,713	1,618	28,474
Total organic carbon [2]	mg/L	3.4	2.7	6.05
Total suspended solids [3]	mg/L	6.8	6	29.3
Sulfate	mg/L	489	31	1,907
Sodium	mg/L	774	416	2,139
Calcium	mg/L	1,079	60	4,177
Magnesium	mg/L	261	71	934
Chloride	mg/L	2,952	623	12,500
Total Silica	mg/L	129	131	134

[1] Menifee Desalter was operated at a recovery of 75 percent.

[2] Of the 16 samples, 6 were less than 2 mg/L or below detection.

[3] Of the 16 samples, 12 were less than 3 mg/L or below detection.

4.3.7 Salt Rejection

Figure 32 shows the EDR salt rejection during Phase 3 testing. This figure shows a fairly constant feed and product TDS and thus a fairly constant salt rejection rate averaging 70 percent, which would be expected during steady and consistent operating conditions. The “noise” shown in the product TDS likely results from the off-spec conductivities measured during polarity reversals, also affecting the salt rejection. If this noise is disregarded, the average salt rejection is seen to be approximately 75 percent.

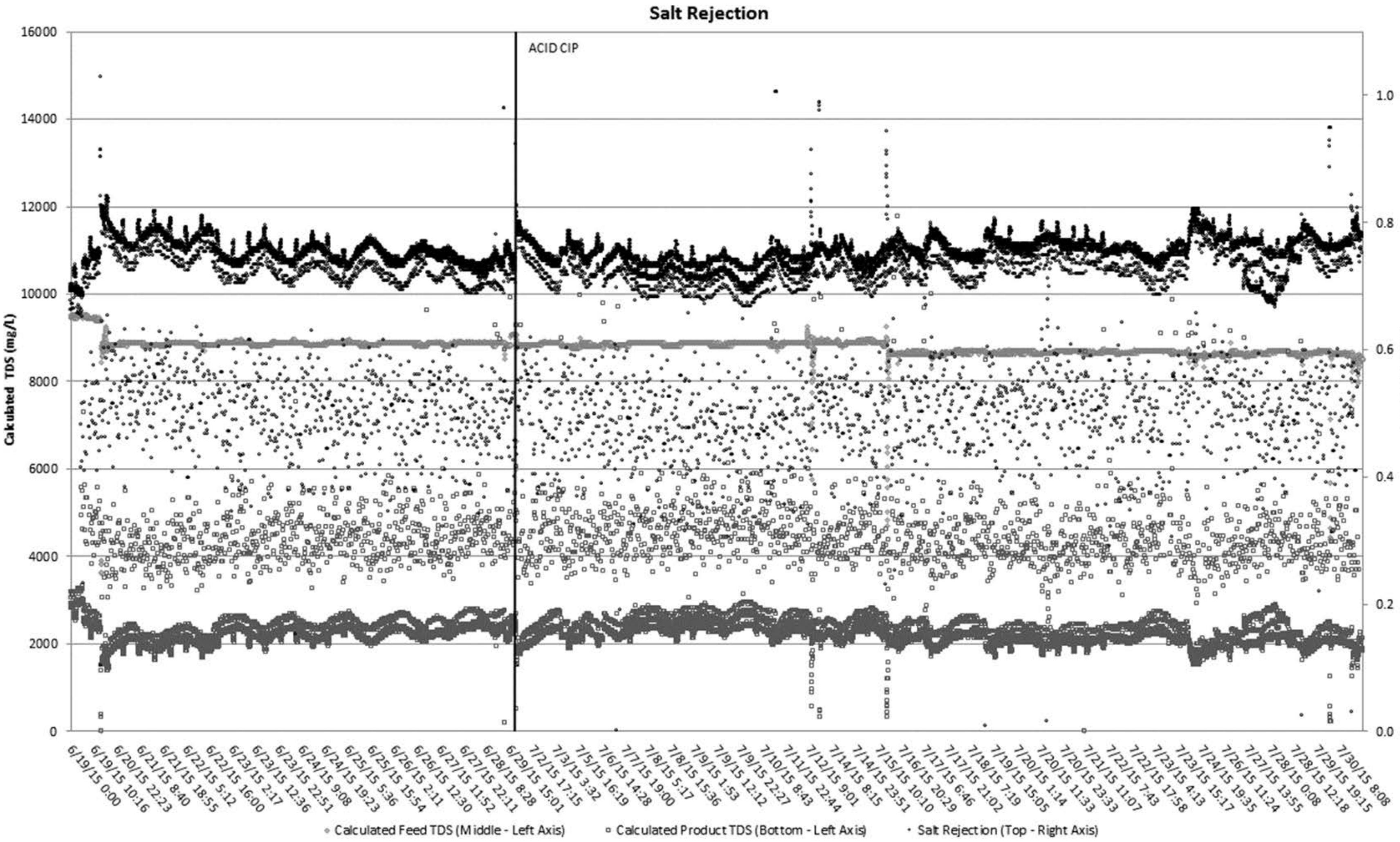


Figure 32. Salt rejection during Phase 3 operations.

4.3.8 Chemical Usage

The chemical usage was obtained from GE. Two chemicals are required in the system; hydrochloric acid for conditioning the electrode compartment and the EDR CIP, and sulfuric acid for conditioning the feed stream to the precipitator. Sulfuric acid is added to the precipitator feed to control calcium carbonate precipitation. Maximum, minimum, and average chemical dosing (50-percent sulfuric acid) observed in the pilot for pH adjustment of the precipitator to approximately 6.3 is presented in Table 12, based on a product water flow rate of 6.4 gpm and a recovery of 80 percent.

Table 12. Sulfuric Acid Usage for AquaSel

Parameter	**Pilot Dosing Rate (liters/hr of 50-percent sulfuric acid)**
Pilot maximum	0.655
Pilot average	0.402
Pilot minimum	0.211
Estimated dose [1]	0.498–0.565

[1] Estimated dose based on pilot results and mass balance modeling.

In addition, hydrochloric acid was used during the operation of the EDR to control the pH in the electrode compartments. Hydrochloric acid was added to the electrode feed on a continuous basis and was increased on a periodic basis for an electrode CIP. The electrode flow in an EDR pilot system does not scale directly to a full-scale system. The pilot chemical usage is the same as one line of full-scale EDR producing 100 gpm. The hydrochloric acid usage during the pilot testing is summarized in Table 13.

Table 13. Hydrochloric Acid Usage for AquaSel

Chemical	**Frequency**	**Injection Rate Per Line of EDR (liters/hr/100 gpm) [1]**	**Daily Consumption Per Line of EDR (liters/day/100 gpm) [1]**
32 percent HCl	Continuous	0.3	7.2
32 percent HCl	Periodic [2]	1.9	5.7
Total 32 percent HCl			12.9

[1] Chemical usage is per 100 gpm of EDR product.

[2] The electrode CIP was performed three times per day for a period of 60 minutes each occurrence.

4.3.9 Other Consumables

Additional consumables for the AquaSel process include cartridge filters and EDR membranes. During the pilot testing, cartridge filters were used in three locations: EDR feed, concentrate return, and precipitator effluent. The cartridge filters on the EDR feed and the concentrate return worked well and would require periodic replacement (approximately every 3 months) based on differential pressure. The cartridge filters on the precipitator effluent required frequent replacement—approximately every 3 to 5 days. This was due to solids being carried over from the precipitator tank and to the precipitation of further solids from the supersaturated solution. An alternative to cartridge filtration would need to be identified for this application at full-scale.

An additional consumable in the EDR process is the EDR membranes themselves. Membrane replacement can be a significant operational expense in EDR, similar to what it is for reverse osmosis. However, EDR membranes are much more durable in concentrate treatment applications. Typical EDR membrane life can vary between 5 and 10 years.

4.4 Phase 3 EDR Concentrate Loop and Precipitator Performance

The precipitator is a critical part of the AquaSel system, and because flow circulates both to and from the precipitator, its performance impacts the performance of the rest of the system. Samples were collected in the streams feeding and returning from the precipitator. However, due to a modification made on the site during Phase 3, the sampling point for the precipitator feed stream was actually sampling a different line. Therefore, unfortunately there is no data for the precipitator feed stream for the 7-week Phase 3 period. However, five data sets that were collected earlier in the testing do include the precipitator feed stream. Table 14 presents the results of these data points.

The calcium sulfate saturation level in the precipitator feed stream was calculated to be 215 percent, based on using the average values presented in Table 14. This indicates that the solution leaving the brine stream was more than twice the theoretical saturation limit with respect to the calcium sulfate, which could be expected. Based on this data, the average conductivity of the precipitator feed was greater than 100,000 µS/cm, which is considerably higher than that reported in Figure 30. Dilution of this stream occurs when the flow mixes with the contents of the precipitator.

Table 14. Precipitator Feed Water Quality (Five Earlier Data Sets)

Parameter	Units	# of Samples	Average	Maximum
Conductivity	µS/cm	5	115,140	191,000
pH	mg/L	5	6.64	7.00
Alkalinity	mg/L	5	744	1,110
Total dissolved solids (TDS)	mg/L	3	([1])	100,000
Nitrate as N	mg/L	5	152	190
Bicarbonate	mg/L	5	906	1,360
Total hardness	mg/L	5	22,860	28,600
Sulfate	mg/L	5	1,790	2,050
Sodium	mg/L	5	3,074	4,060
Potassium	mg/L	5	120	160
Total phosphorus	mg/L	5	1.04	1.70
Calcium	mg/L	5	6,348	8,010
Total Iron	mg/L	1	290	- -
Magnesium	mg/L	5	1,700	2,080
Chloride	mg/L	5	19,500	24,200
Fluoride	mg/L	5	1.01	2.50
Total silica	mg/L	5	140	160
Reactive silica	mg/L	5	128	150

[1] Insufficient data.

Table 15 presents the results of the precipitator effluent water quality. This stream returns to the EDR stack as the brine loop after being filtered to remove any residual suspended solids. As shown in Table 15, the suspended solids in the precipitator effluent stream had an average concentration of about 30 mg/L. Based on the flow rate of 5.2 gpm for the precipitator effluent, this represents 0.08 pounds per hour (lb/hr) of solids.

Table 15. Precipitator Effluent Water Quality

Parameter	Units	# of Samples	Average	Maximum
Conductivity	µS/cm	63,088	42,770	121,199
Turbidity	NTU	16	1.67	7.20
Alkalinity	mg/L	19	1,093	1,500
Total dissolved solids (TDS)	mg/L	19	28,474	39,000
Total organic carbon	mg/L	16	6.05	8.50
Total suspended solids	mg/L	0	29.3	63.0
Nitrate as N	mg/L	19	84.0	110.0
Bicarbonate	mg/L	19	1,332	1,820
Total hardness	mg/L	19	14,274	17,263
Barium	µg/L	16	1,507	3,500
Strontium	mg/L	16	21.6	31.9
Sulfate	mg/L	19	1,907	2,490
Sodium	mg/L	19	2,139	2,710
Ammonia as N	mg/L	16	0.6	1
Potassium	mg/L	19	87.4	110.0
Total phosphorus	mg/L	19	1.6	3.6
Calcium	mg/L	19	4,177 [1]	5,110
Total iron	mg/L	17	152.8	500.0
Magnesium	mg/L	19	934	1,100
Manganese	µg/L	16	0	0
Chloride	mg/L	19	12,500	13,200
Fluoride	mg/L	19	1.4	1.5
Boron	µg/L	16	385	740
Aluminum	µg/L	16	35.0	65.0
Arsenic	µg/L	16	35.1	44.0
Copper	µg/L	16	12.1	30.0
Selenium	µg/L	16	54.4	72.0
Total silica	mg/L	19	134	170
Reactive silica	mg/L	19	141	150

[1] A minimum calcium value of 1,100 mg/L occurred on June 24, 2015. The value was included in the calculation of the average, but is the only value that is below 3,500 mg/L and may be an error. See Figure 33.

Figure 33 presents variations of bicarbonate, calcium, sulfate, and TDS concentrations in the precipitator effluent for Phase 3, as well as for one earlier data point on May 18. The scale for bicarbonate, calcium, and sulfate concentrations is shown on the left side of the figure; the scale for TDS is on the right.

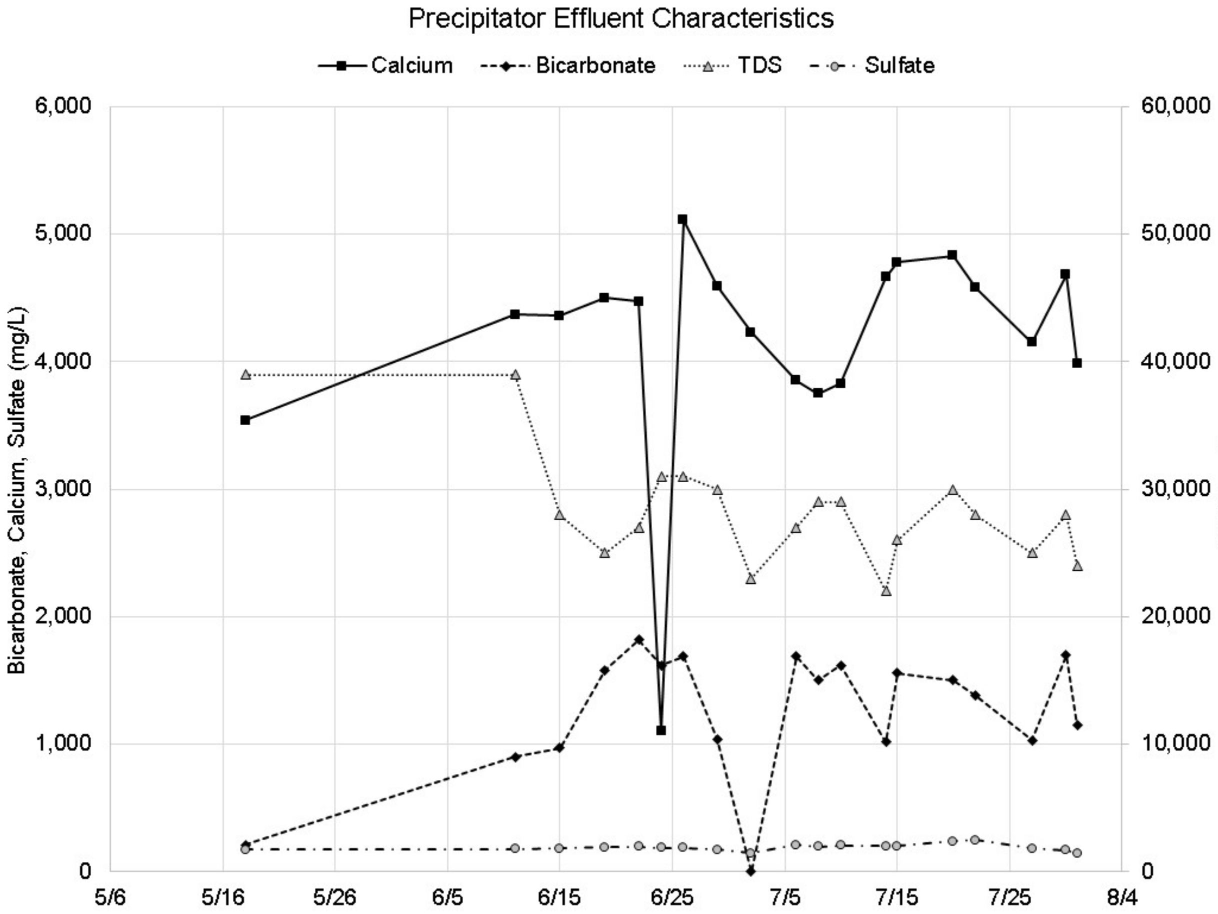


Figure 33. Variation of selected parameters in precipitator effluent with time.

For the period of June 15 to July 31, 2015, calcium typically varied between 4,000 and 5,000 mg/L. There was one significant outlier on June 24, which appears to be a sampling error. Bicarbonate was typically between 1,000 and 1,800 mg/L, and sulfate was between 1,700 and 2,400 mg/L. The TDS varied between 20,000 and 30,000 mg/L.

Based on the average values in Table 15, the calcium sulfate saturation concentration in the precipitator effluent was calculated to be 200 percent. Calculations show that to bring this stream to equilibrium with respect to calcium sulfate (that is around 100 percent saturation), between 3 and 4 lb/hour of gypsum would need to precipitate.

4.4.1 Purpose of Precipitator

The purpose of the precipitator is to precipitate calcium sulfate, which can be removed from the system, so that a less saturated solution can be returned to the EDR stack to collect additional ions. Figure 34 depicts diagrammatically what occurs in the precipitator. The feed stream to the precipitator will be supersaturated with calcium sulfate. Based on the average water quality presented above, the feed stream saturation was calculated as 215 percent. If given enough time, and in the presence of other seed crystals, the saturation level of the feed stream would drop and could follow the dotted line on Figure 34, which indicates the theoretical curve for percent saturation level. If equilibrium were to be achieved, the percent saturation would reach 100 percent. In a single mixed reactor, however, with limited detention time this is not possible and the saturation level will be at some point above 100 percent. The potential amount of solids that could precipitate in the precipitator is represented by the percent saturation level of the feed stream. The actual amount that precipitates and can be extracted from the system is represented by the difference between the feed saturation level and the precipitator effluent saturation level. The precipitator effluent saturation level represents dissolved solids that are returned to the system.

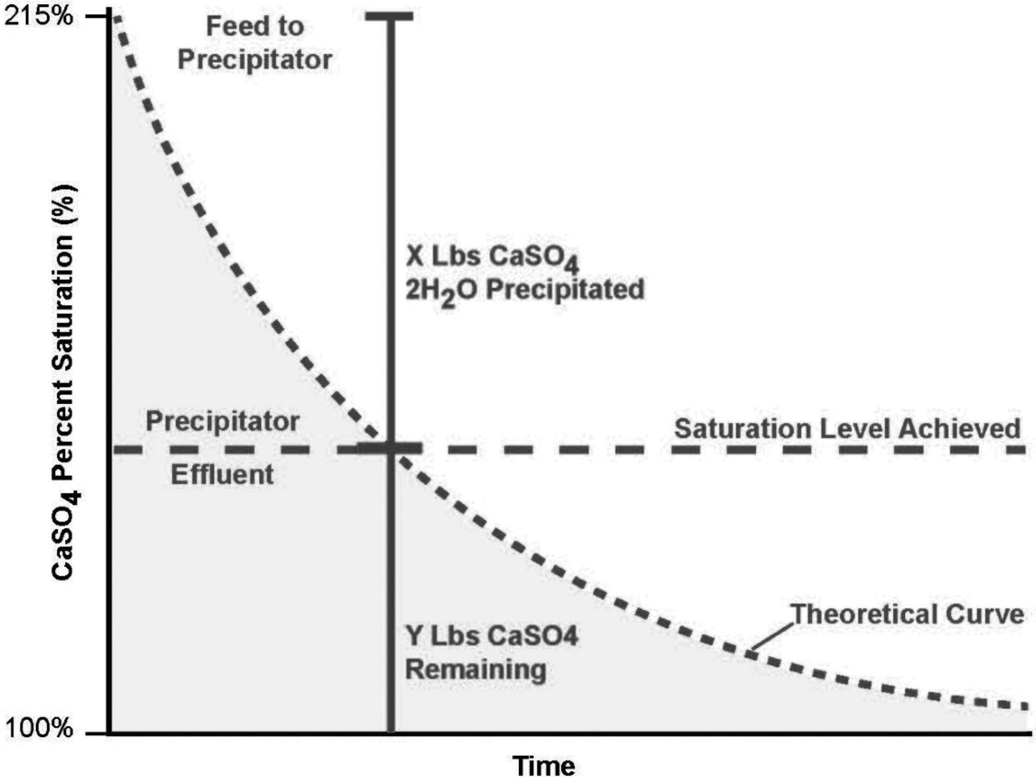


Figure 34. Diagrammatic representation of precipitator operation.

As mentioned in the previous section, the average water quality of the precipitator effluent indicated a saturation level of around 200 percent. The difference between 215 percent in the feed stream and the 200 percent in the effluent does not indicate substantial solids removal; and as mentioned, there was potential to precipitate an additional 3 to 4 lb/hr of gypsum.

4.4.2 Solids Production

The waste stream from the AquaSel system was blown down periodically from a recirculating solids stream on the precipitator tank. The blowdown represented discharge of both excess solids and liquid from the system and was used to control the mass balances of the liquid and solid phases. This is discussed further in the following section. Table 16 presents the analytical data for the liquid phase of the blowdown stream from the precipitator tank. To obtain the results for the liquid phase blowdown, the waste blowdown stream was filtered and then analyzed for the parameters shown in the table.

Table 16. AquaSel Blowdown Effluent Water Quality

Parameter	Units	# of Samples	Average	Maximum
Conductivity	µS/cm	63,088	42,750	188,175
pH		16	7.1	7.7
Alkalinity	mg/L	0	([1])	([1])
Total dissolved solids (TDS)	mg/L	16	30,103	84,000
Total organic carbon	mg/L	16	5.98	8.70
Nitrate as N	mg/L	19	97	130
Barium	µg/L	16	3.8	4.6
Strontium	mg/L	16	30	33
Sulfate [2]	mg/L	19	1,651	2,380
Sodium	mg/L	19	2,403	2,594
Potassium	mg/L	19	81	86
Calcium	mg/L	19	4,808	5,208
Total iron	mg/L	17	217	525
Magnesium	mg/L	19	1,090	1,169
Manganese	µg/L	16	< 250	- -
Chloride	mg/L	19	12,688	14,000
Boron	µg/L	16	358	774
Aluminum	µg/L	16	180	210
Arsenic	µg/L	16	41	44
Copper	µg/L	16	38	54
Selenium	µg/L	16	70	78
Total silica	mg/L	19	132	138

[1] Not measured.

[2] One value of 15,607 mg/L reported for sulfate was removed from the dataset as it is likely a sampling error.

Of the major ionic species, sodium, magnesium, and chloride showed fairly consistent concentrations. However, TDS, sulfate, and calcium showed some significant variation. (See Figure 33.)

The calcium sulfate saturation level on the blowdown stream was calculated to be around 183 percent based on the average values in Table 16. This is lower than the value obtained for the precipitator effluent (Table 15 above), probably because the sample bottles contained a lot more solids than calculated, and precipitation could have continued after sampling.

Figure 35 shows the variation in calcium and sulfate concentrations in the blowdown stream. The error bars indicated one standard deviation. Most results were within one standard deviation, but the variation in sulfate concentration was greater than that of calcium.

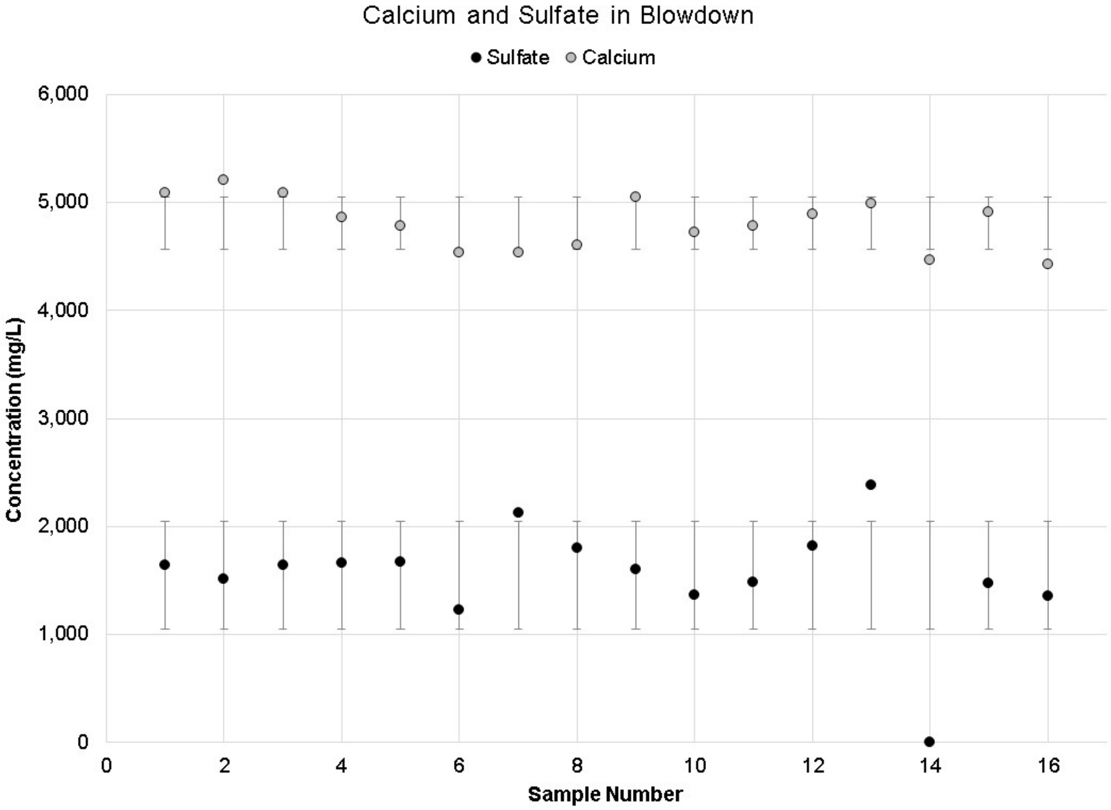


Figure 35. Variation of calcium and sulfate concentrations in the blowdown stream.

The solids phase from the waste blowdown stream was also analyzed. The solids were separated, digested, and then analyzed for metals concentration. Most of the tested metal species had a result of less than 0.01 percent of the sample and are therefore assumed to contribute an insignificant quantity to the solids. Table 17 shows the metals that were detected in higher concentrations. The units are percent of the total solids present. As expected, calcium had the greatest percentage.

Table 17. Metals in AquaSel Blowdown Solids

Parameter	Units	# of Samples	Average	Maximum
Barium	Percent	16	0.01	0.03
Strontium	Percent	16	0.06	0.16
Sodium	Percent	16	1.24	11.9
Potassium	Percent	16	0.06	0.50
Calcium	Percent	16	28.2	29.3
Magnesium	Percent	19	0.58	5.39
Silica (SiO_2)	Percent	16	0.58	0.67

The solids blowdown rate was reported as 0.1 gpm plus the precipitator overflow rate of 1.5 gpm, for a total of 1.6 gpm. The solids blowdown stream had an average solids concentration of 11.2 percent, but this was diluted when it was mixed with the precipitator overflow stream. Figure 36 shows the variation in solids concentration of the blowdown stream with time. As shown, there was a lot of variability, and the general trend was a reduction in solids concentration with time. It appears from the data that the performance of the precipitator was not stable. At an average solids concentration of 11.2 percent by weight and a blowdown rate of 0.1 gpm, the amount of solids discharged from the system averaged 5.6 lb/hr.

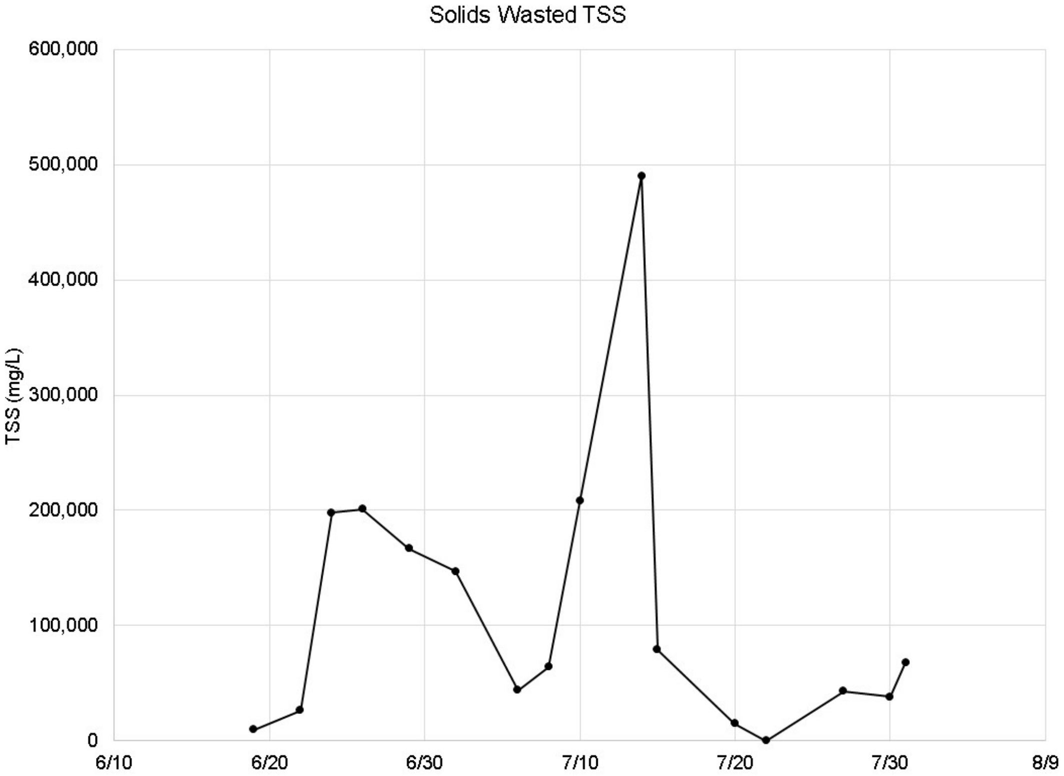


Figure 36. Solids concentration in waste blowdown stream.

Five sets of data were available to assess the change in water quality across the precipitator. This data was collected earlier in the pilot plant study and not during Phase 3, and is presented in Table 18. Only three sets of calcium data were available. The average change in calcium concentration was 422 mg/L. The difference between the calcium and sulfate molarity change indicates that the change in sulfate controls the precipitation of calcium sulfate (gypsum). The figures in Table 16 suggest that around 2.6 lb/hr of gypsum precipitated.

The higher loss of calcium suggested that another precipitate of calcium was also forming, most likely calcium carbonate. Accounting for the loss of calcium with gypsum formation leaves calcium that could form about 3.2 lb/hr of calcium carbonate.

Table 18. Change in Calcium and Sulfate Concentration Across Precipitator

Sample Date	**Calcium (mg/L)**			**Sulfate (mg/L)**		
	IN	**OUT**	**Change**	**IN**	**OUT**	**Change**
April 24, 2015	N/A	N/A	- -	1,810	1,530	390
April 29, 2015	7,260	6,360	900	1,580	1,330	360
May 4, 2015	8,010	7,420	590	1,720	1,430	400
May 11, 2015	7,130	6,520	610	1,790	1,370	530
May 18, 2015	N/A	N/A	- -	2,050	1,730	430
Average	-	-	700	-	-	422

N/A = Not applicable.

4.4.3 Solids Mass Balance

The section above indicated that 2.6 lb/hr of gypsum solids were formed in the precipitator between April and May. Of that, we know that 0.08 lb/hr was returned to the EDR process in the solids that overflowed the precipitator tank. This material was captured in the candle filter and cartridge filters. For the solids balance to be maintained, the remaining solids must have been wasted from the system. Most of these solids would be discharged with the blowdown stream, and as mentioned above, this appeared to be an average of 5.6 lb/hr across Phase 3.

5.6 lb/hr is more than the amount of gypsum that was calculated to form, but is approximately the correct number if about 3.2 lb/hr of calcium carbonate is included with the solids. The solids would then be a mixture of gypsum and calcite ($CaCO_3$). Based on the above values, the ratio would be 43 percent gypsum and 57 percent calcite.

If this ratio is correct, then calcium would represent about 32.8 percent of the solids content. The laboratory analysis (Table 19) showed the percentage to range from 26.4 to 29.3, with an average of 28.2 percent. A pure gypsum sample would have a calcium concentration of 23.3 percent. However, the solids samples were heated to about 105 °C prior to processing, which would have driven off most of the water of hydration and converted the gypsum to the hemihydrate form ($CaSO_4 \cdot ½H_2O$). In this case the calcium concentration would be 27.5 percent. The higher average concentration of calcium concentration would be 27.5 percent. The higher average concentration of calcium indicates that some other calcium precipitates were present in the solids.

Table 19. Weight of Dried Cartridge Filters

Sample Date	Cartridge Filter Number and Dried Weight, lbs			Operating Period
	1	2	3	
July 17, 2015	11	12	12	4 days
July 21, 2015	9.5	9.0	11	2 days
July 23, 2015	9.0	8.0	8.5	

4.4.4 Solids Captured in Cartridge Filters

As mentioned above, most of the precipitator product stream (overflow) was returned to the EDR stack concentrate loop after filtration. The suspended solids in the precipitator overflow stream (0.08 lb/hr on average) were captured by the cartridge filters. These filters required changing every 3 or 4 days due to the buildup of solids. Based on the dried weight of the used cartridge filters, an average weight gain per hour could be calculated. Data collected by EMWD operators is shown in Table 19. Using this data, the average amount of solids captured by the cartridge filters was calculated to be 0.48 lb/hr (range was 0.36–0.6 lb/h). This is significantly more than the suspended solids that were present in the precipitator overflow stream, indicating that more than 80 percent of the solids captured in the cartridge filters actually precipitated there. This is not surprising considering that the precipitator overflow stream was supersaturated with calcium sulfate (around 200 percent of saturation on average), by our calculations.

4.4.5 Component Mass Balance

A mass balance across the AquaSel pilot system was carried out in order to check whether there was a balance between measured parameters. There were no suspended solids entering the system, just dissolved species. The mass balance was calculated for six parameters: calcium, sulfate, sodium, chloride, nitrate, and total silica. Because sodium, chloride, and nitrate species do not precipitate at the concentrations encountered in this system, they were used to check the accuracy of the balance. Silica was also used as a parameter for the mass balance, since it should pass through the EDR system unchanged.

Overall, the system contains one feed stream, the raw water, and three discharge streams: the EDR product water, the concentrate blowdown, and the precipitator blowdown. There is one additional source of sulfate in the system: the sulfuric acid addition. The concentrate and precipitator blowdown streams contain both solids and liquids.

Two mass balances were carried out: one that includes all water quality data obtained from the laboratory, and a second one in which “outlier” data was removed from the EDR product quality data.

Figure 37 shows a comparison of the EDR product water stream with all the data (top) and excluding the outliers (bottom). The spikes in the data of several parameters are clearly seen in the top part of the figure. It should be noted that, for those parameters showing spikes, the spikes did not all occur on the same sample date. It is not clear why some parameters showed spikes in one sample and different parameters showed spikes in another sample.

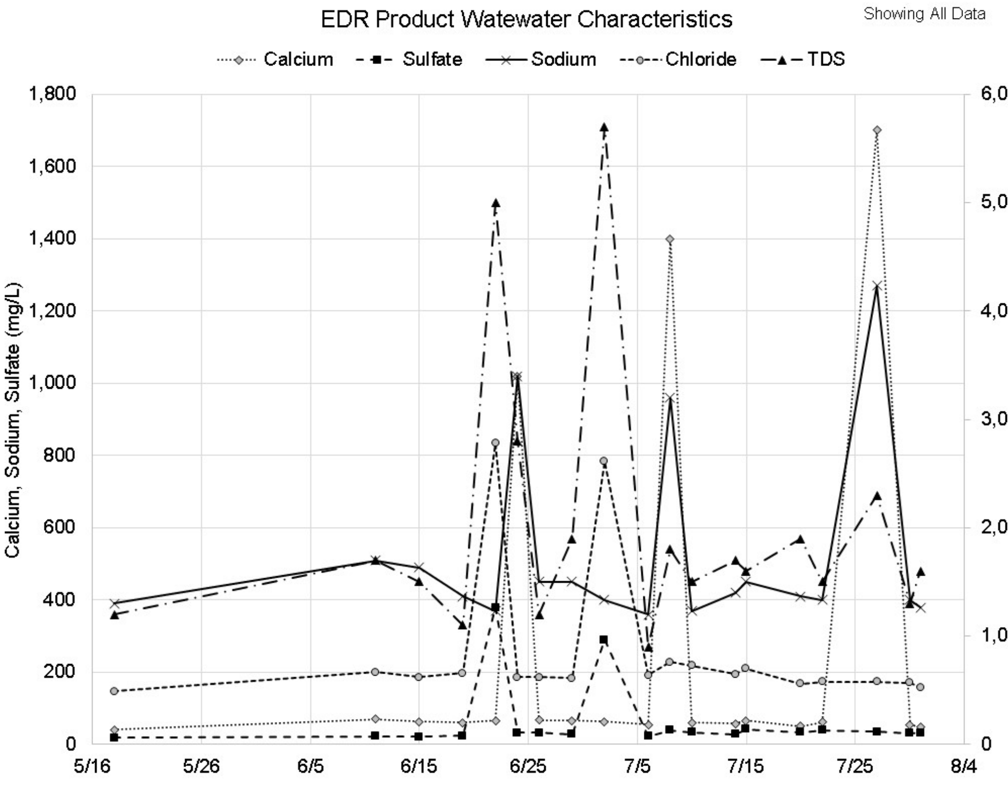
Table 20 presents the average concentrations of calcium, sulfate, sodium, chloride, nitrate, and silica in the four streams mentioned above. The data for the EDR product stream is divided into two. The left-hand column shows the average of all the data including the outliers, and the right-hand column shows the average values without the outliers. The silica concentration did not show any peak values. The two blowdown streams have been combined, since the dissolved parameters in both streams will be the same. Sulfuric acid addition adds 110 mg/L of sulfate to the precipitator, for a total of 0.37 lb/hr.

Table 20. Average Mass Balance Concentrations

Parameter	**Units**	**Raw Feed**	**EDR Product**		**Concentrate and Precipitator Blowdown** [1]
			With All Data	**Excluding Outliers**	
Flow rate	gpm	8.0	6.4	6.4	1.5 + 0.1 = 1.6
Sulfate	mg/L	489	63	30	2,523
Sodium	mg/L	774	522	416	2,403
Calcium	mg/L	1,079	267	60	4,808
Chloride	mg/L	2,953	841	841	12,688
Nitrate as N	mg/L	22	6.1	4.7	97
Total silica	mg/L	129	131	131	132

[1] Sulfuric acid addition at 225 mg/L adds 110 mg/L of sulfate to the system.

Using the concentrations and flows in Table 20, the mass of each ion into and out of the system was calculated. These values are presented in Tables 21 and 22. Table 22 includes calculations using the EDR product stream data that excludes the outlier data. The difference between the “Total In” and the “Total Out” was calculated and expressed as a percentage of “Total In.”



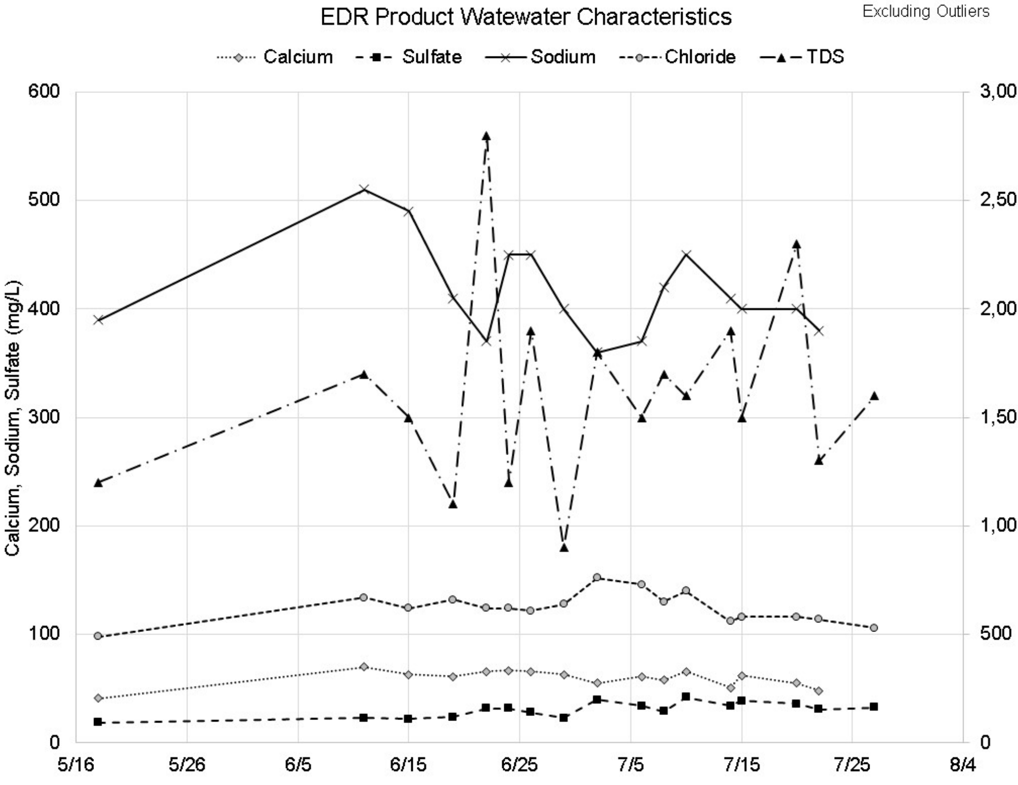


Figure 37. Comparison of EDR product quality with all data (above) and without "outliers" (below).

As can be seen from the percentage difference column in Table 21, the closest balance was obtained with total silica (2 percent). For the other totally dissolved species such as sodium, chloride, and nitrate, the balance was within 16 percent, 9 percent, and 10 percent respectively. The sulfate balance showed that sulfate was lost from the system, which was as expected, but only 5 percent in this case, which would not represent much in terms of gypsum precipitation. Calcium showed a gain of 9 percent.

Table 21. Mass Balance Calculations for AquaSel (All EDR Product Data)

Parameter	Units	Total In	Total Out	Difference (percent)
Sulfate	lb/hr	2.34 [1]	2.23	–5
Sodium	lb/hr	3.11	3.60	+16
Calcium	lb/hr	4.32	4.72	+9
Chloride	lb/hr	11.8	12.9	+9
Nitrate as N	lb/hr	0.088	0.097	+10
Total silica	lb/hr	0.516	0.525	+2

[1] Sulfuric acid addition at 225 mg/L adds 110 mg/L of sulfate to the system.

The values presented in Table 22, on the other hand, show a different result. Here, the fully dissolved species of sodium, chloride, nitrate, and silica are all within 5 percent, indicating good agreement with the mass balance. Sulfate showed a decrease of 9 percent and calcium a decrease of 6 percent, indicating that both of these ions were lost from the system.

Table 22. Mass Balance Calculations for AquaSel (Excluding Outliers)

Parameter	Units	Total In	Total Out	Difference (percent)
Sulfate	lb/hr	2.34 [1]	2.12	–9
Sodium	lb/hr	3.11	3.22	+4
Calcium	lb/hr	4.32	4.04	–6
Chloride	lb/hr	11.8	12.2	+3
Nitrate as N	lb/hr	0.088	0.092	+5
Total silica	lb/hr	0.516	0.525	+2

[1] Sulfuric acid addition at 225 mg/L adds 110 mg/L of sulfate to the system.

The change in sulfate mass balance shown in Table 22 suggests a loss of 0.22 lb/hr. The change in sulfate concentration across the precipitator, presented earlier, suggests that around 2.57 lb/hr of gypsum is produced. This would account for about 1.4 lb/hr of sulfate, which is significantly more than the 9-percent difference shown in Table 22. Closer inspection of the waste blowdown stream data shows one very high value of sulfate. (See Table 18.) By eliminating this data point, the change in sulfate mass balance increased to 47 percent, which would account for about 1.63 lb/hr of gypsum produced. This modification of the data does not fully account for the observed change in sulfate concentration across the precipitator, but accounts for about 63 percent of it.

The differences observed in the sulfate mass balance suggest a few things:

- Either the system was not stable, resulting in variable data; or
- Precipitation continued in the sample bottles, which skewed the data; or
- Both of the above occurred.

The data was analyzed further to clarify the reason(s) for the differences in the observed mass balance data. In this case a system-wide balance was performed for both calcium and sulfate for each day that sample data was available. The mass balance is described below for calcium and sulfate:

Calcium IN to the system in the Menifee Brine
= Calcium OUT in the product stream
+ Calcium OUT in the liquid of the blowdown stream
+ Calcium OUT in the solids of the blowdown stream
+ Calcium OUT in the Cartridge Filters
+ Calcium ACCUMULATION within the system.

Sulfate IN to the system in the Menifee Brine
= Sulfate OUT in the product stream
+ Sulfate OUT in the liquid of the blowdown stream
+ Sulfate OUT in the solids of the blowdown stream
+ Sulfate OUT in the Cartridge Filters
– Sulfate IN with sulfuric acid
+ Sulfate ACCUMULATION within the system.

A schematic for the sulfate balance is shown in Figure 38.

Table 23 shows the mass balance for calcium for each available sample during Phase 3. The mass of calcium removed from the system in the cartridge filters was estimated from the weight of the cartridge filters and an assumption that the solids were all calcium sulfate. A constant average value of 0.11 lb/hr of calcium was assumed for each data set. This is a relatively small amount compared with the total amount of calcium entering the system, so it should not skew the data severely if the estimate is inaccurate.

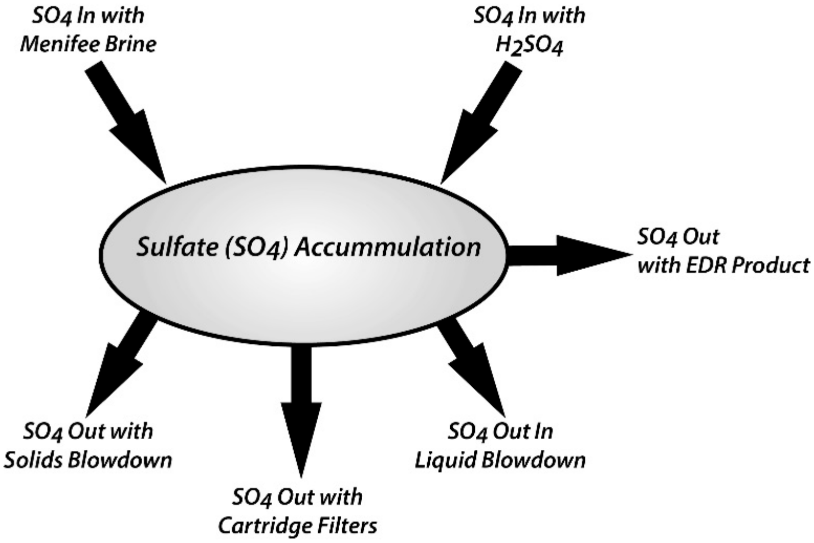


Figure 38. Schematic of mass balance for sulfate.

Table 23. Calcium Mass Balance by Sample Date for AquaSel

Sample Number	Sample Date	Total Calcium IN (lb/hr)	Total Calcium OUT (lb/hr)	Accumulation = IN – OUT (lb/hr)	Difference = Accumulation / IN (percent)
1	6/19/2015	4.41	4.52	–0.11	–2.49
2	6/22/2015	4.08	4.85	–0.77	–18.87
3	6/24/2015	3.97	4.39	–0.42	–10.58
4	6/26/2015	4.65	7.03	–2.38	–51.18
5	6/29/2015	4.70	6.59	–1.90	–40.43
6	7/2/2015	4.72	5.89	–1.17	–24.79
7	7/6/2015	4.32	4.52	–0.20	–4.63
8	7/8/2015	3.44	4.90	–1.46	–42.44
9	7/10/2015	3.48	7.36	–3.88	–111.49
10	7/14/2015	4.45	10.83	–6.37	–143.15
11	7/15/2015	4.45	5.20	–0.75	–16.85
12	7/20/2015	4.41	4.19	0.22	4.99
13	7/22/2015	4.48	4.30	0.18	4.02
14	7/27/2015	4.41	4.50	–0.09	–2.04
15	7/30/2015	4.41	4.78	–0.37	–8.39
16	7/31/2015	4.26	4.81	–0.55	–12.91

Figure 39 presents the total calcium put into the system and compares it with the total calcium that came out of the system for each data point shown in Table 23. As can be seen, there are 10 data points for which the in and out data are relatively close. The other six data points (#'s 4, 5, 6, 8, 9, and 10) are way off. In almost all cases the data indicated that there was more calcium leaving the system than entering, suggesting that there was not a large accumulation of calcium in the system during Phase 3.

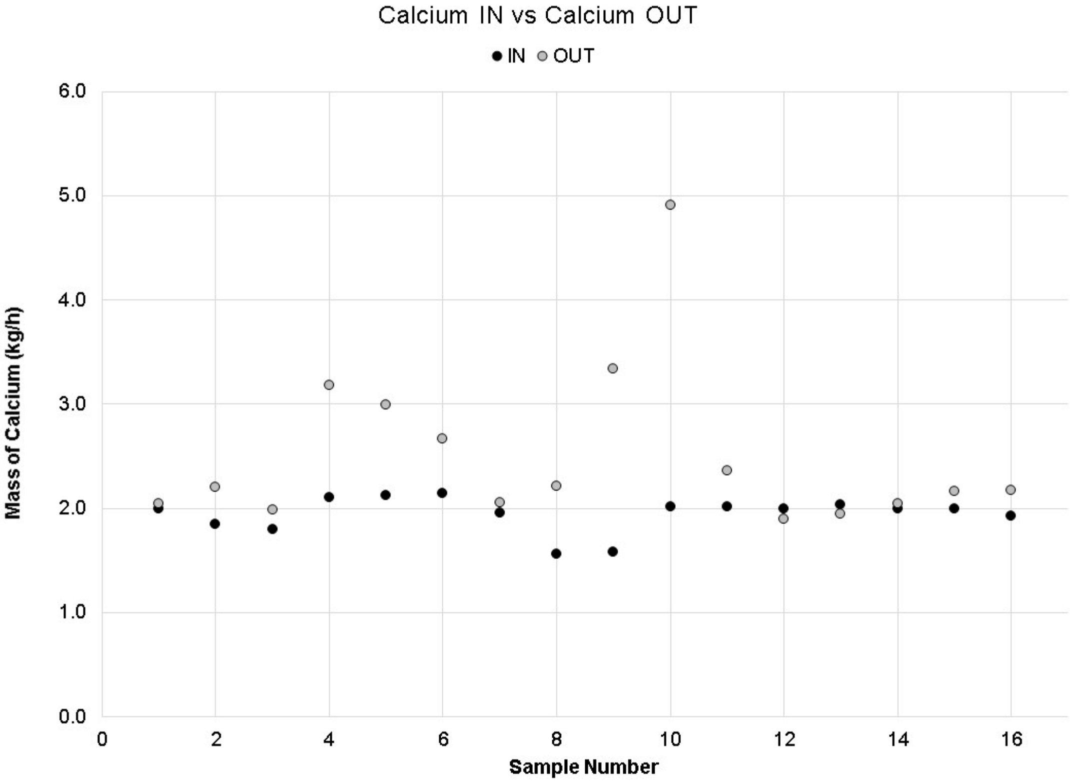


Figure 39. Calcium mass balance across AquaSel system.

Similarly, Figure 40 presents the mass balance for sulfate into and out of the AquaSel system for each of the same data points. An average of 0.37 lb/hr of sulfate from acetic acid addition was used. In this case there was no direct measurement of sulfate available in the precipitated solids, like there was for calcium. The sulfate associated with the calcium depends on the amount of gypsum and other calcium salts that precipitated. If the solids formed were pure gypsum then the molar contribution by sulfate would be 100 percent of the calcium. Based on the previous discussion (Section 4.4.3), it appears that the solids were not pure gypsum. However, because the ratio of gypsum to other calcium precipitates is not known, an estimated percentage was used based on the closeness of the sulfate mass balance. To determine the estimate, the percentage was adjusted on a trial and error basis and the "sulfate out" total was observed in relation to the "sulfate in" total. Figure 40 is based on the estimate that 40 percent of the calcium in the precipitated solids was bound with sulfate. This resulted in the "closest" mass balance, with six data sets almost matching and four others fairly close, but indicating some accumulation of sulfate in the system.

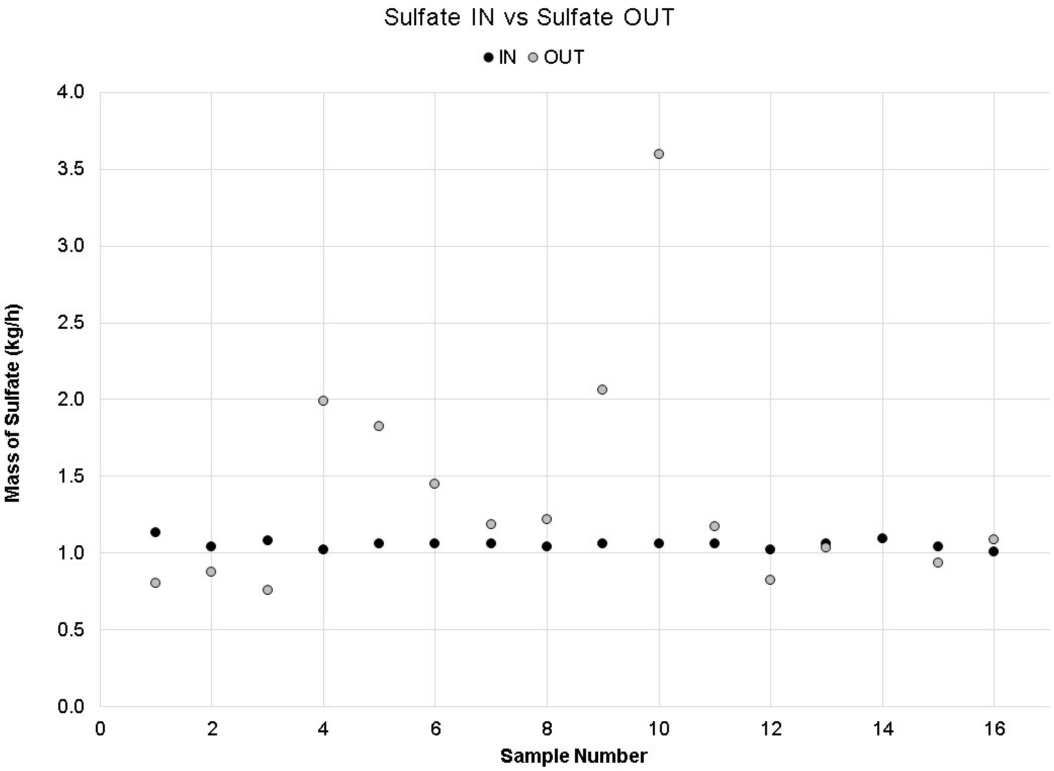


Figure 40. Sulfate mass balance across AquaSel system based on 40 percent associated with calcium precipitate.

Overall though, similar to the case for calcium, there is a lot of scatter in the data and there is a wide deviation from the mass balance between data points 4 (June 26th) and 10 (July 14th), with the period of stability for data points 7 (July 6th) and 8 (July 8th). This trend is the same for the calcium data, and could indicate a period of instability during the operation of the precipitator.

The analysis of the individual data points has shown that there are inconsistencies in the data and that averaging across the Phase 3 data set to determine the mass balance could lead to inaccuracies and cause difficulty in obtaining a mass balance, as was the case here.

5.0 PRELIMINARY COST ANALYSIS

A preliminary capital and operation and maintenance (O&M) cost estimate was developed for the AquaSel process based on the pilot data. The cost presented represents the planning level capital and O&M costs associated with an AquaSel process designed to treat 2.4 mgd of RO concentrate.

5.1 Cost Estimate Assumptions

Several baseline cost assumptions were required to complete the cost estimate for the concentrate treatment process. These assumptions include O&M factors, such as cost for power, chemicals, etc., and estimates of the process capital cost based on past projects and vendor quotes. At the time of this cost estimate (October 2015), the Los Angeles-based *Engineering News-Record* construction cost index was 10,981.

The capital cost estimate presented below represents a Class 5 estimate as defined by the Association for the Advancement of Cost Engineering International.

5.1.1 Capital Cost Assumptions

Several planning-level cost assumptions were made based on vendor quotes for equipment and rule-of-thumb parameters for membrane treatment costs. A summary of capital cost assumptions is shown in Table 24.

Table 24. Capital Cost Assumptions

Parameter	Value	Units
Yard piping [1]	3	Percent
Interconnecting pipework [1]	14	Percent
Electrical and instrumentation [1]	18	Percent
Building costs	150–200	$/sq ft
Contingency	25	Percent
Contractor general conditions	5	Percent
Contractor overhead and profit	10	Percent
Bid market allowance [2]	0	Percent
Sales tax [3]	8	Percent
Engineering	10	Percent
Construction management	10	Percent
Administration, environmental and permits	8	Percent

[1] As a percentage of direct equipment costs.

[2] Biding conditions were not taking into consideration.

[3] Assumed that this applies to 50 percent of the equipment costs and 25 percent of the yard piping costs.

5.1.2 Operation and Maintenance Cost Assumptions

A list of the assumptions used to determine the O&M costs of the AquaSel system is presented in Table 25.

Table 25. O&M Cost Assumptions

Parameter	Units	Value	Estimated Consumption
Hydrochloric acid (32 percent)	$/gal	$4.50	55 gal/day [1]
Sulfuric acid (50 percent)	$/gal	$3.30	317 gal/day
Cartridge filters	$/filter	$300	Monthly changes
EDR membrane replacement	$/1,000 gal	$0.20	- -
Sludge disposal	$/ton	$50	Based on biosolids disposal cost
Electrical power	$/kWh	$0.13	9.2 kWh/kgal

[1] Excludes acid used for CIPs – about 4,000 gal/month.

5.2 Cost Estimate for 2.4-mgd Process Unit

Using the assumptions and approach stated above, an overall project cost was developed.

5.2.1 Capital Cost Estimate

The planning level Class 5 capital cost estimate for a treatment train to treat 2.4 mgd or 1,667 gpm of Menifee and Perris I Desalter concentrate at a recovery of 80 percent is summarized in Table 26. The project cost estimate is $40.1 million. This estimate includes chemical systems, process equipment, solids handling, pumping, and other ancillary facilities.

The treatment system would be made up of four 400-gpm modules. Figure 41 shows an example of what a single four-stage 400-gpm AquaSel module would look like. As shown, each module would include 16 EDR stacks. The precipitator tanks are not shown. The modules would be arranged within a building and would also include precipitator tanks — one or two per 400-gpm module.

5.2.2 Operation and Maintenance Cost Estimate

The planning level O&M costs for the treatment facilities are summarized in Table 27. The total annual O&M cost is estimated to be approximately $3.1 million. This amount includes electrical costs, chemical costs, membrane replacement, brine disposal, solids disposal and an allowance for providing one additional operator.

Table 26. Capital Cost Estimate - 2.4-mgd AquaSel System

Component Description	Cost [1,2]
Site civil and yard work	2,095,459
AquaSel system [3]	8,966,000
Dewatering / loading facility	2,310,000
Chemicals, product storage and transfer	1,050,231
Equipment installation	4,034,700
Buildings	1,250,000
Electrical, instrumentation, and controls	2,218,721
Contractor general conditions, 5 percent	1,096,256
Contractor overhead and profit, 10 percent	2,192,511
Bid market allowance, 0 percent	0
Sales tax @ 8.0 percent [4]	666,312
Contingency, 25 percent	5,481,278
Construction cost	**31,361,467**
Engineering, 10 percent	3,136,147
Construction management, 10 percent	3,136,147
Administration, environmental and permits, 8 percent	2,508,917
Total estimated project cost	**40,142,678**

[1] *Engineering News-Record* construction cost index for Los Angeles (October 2015, 10,981).

[2] Sub-total of direct costs which includes all equipment, buildings and installation is $21,925,111

[3] Estimate provided by GE

[4] Calculated assuming 50 percent of equipment costs and 25 percent of yard piping are taxable.

Table 27. Operation and Maintenance Cost Estimate — 2.4-mgd AquaSel System

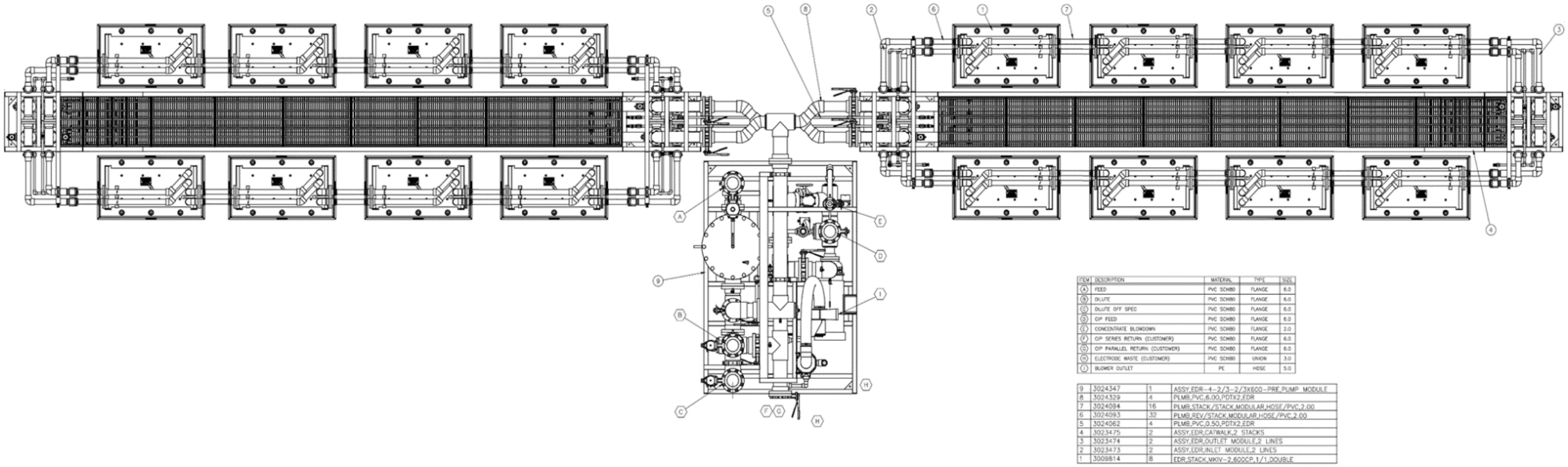
Parameter	Value	Units	Source
Equipment O&M costs [1]	123,262	$/yr	See note
Chemical costs	256,686	$/yr	Pilot testing
Solids disposal costs	415,917	$/yr	$50/ton [2]
Brine disposal costs	493,880	$/yr	Based on Inland Empire Brine Line
EDR membrane replacement costs	175,347	$/yr	Estimate
Cartridge filters	43,200	$/yr	Pilot/estimate
Power costs	1,433,889	$/yr	Pilot testing [3]
Labor and staffing costs allowance [4]	115,000	$/yr	Estimate
Total cost	**3,057,182**	$/yr	

[1] Assumed to be 1 percent of the equipment cost.

[2] Based on 17 wet t/d of solids produced.

[3] Based on 9.2 kWh/kgal estimate provided by GE.

[4] Allowance for one additional operator.



ITEM	DESCRIPTION	MATERIAL	TYPE	SIZE
Ⓐ	FEED	PVC SCH80	FLANGE	6.0
Ⓑ	DILUTE	PVC SCH80	FLANGE	6.0
Ⓒ	DILUTE OFF SPEC	PVC SCH80	FLANGE	6.0
Ⓓ	CIP FEED	PVC SCH80	FLANGE	6.0
Ⓔ	CONCENTRATE BLOWDOWN	PVC SCH80	FLANGE	2.0
Ⓕ	CIP SERIES RETURN (CUSTOMER)	PVC SCH80	FLANGE	6.0
Ⓖ	CIP PARALLEL RETURN (CUSTOMER)	PVC SCH80	FLANGE	6.0
Ⓗ	ELECTRODE WASTE (CUSTOMER)	PVC SCH80	UNION	3.0
Ⓘ	BLOWER OUTLET	PE	HOSE	5.0

9	3024347	1	ASSY,EDR-4-2/3-2/3X600-PRE,PUMP MODULE
8	3024329	4	PLMB,PVC,6.00,PDTX2,EDR
7	3024094	16	PLMB,STACK/STACK,MODULAR,HOSE/PVC,2.00
6	3024093	32	PLMB,REV/STACK,MODULAR,HOSE/PVC,2.00
5	3024062	4	PLMB,PVC,0.50,PDTX2,EDR
4	3023475	2	ASSY,EDR,CATWALK,2 STACKS
3	3023474	2	ASSY,EDR,OUTLET MODULE,2 LINES
2	3023473	2	ASSY,EDR,INLET MODULE,2 LINES
1	3009814	8	EDR,STACK,MKIV-2,600CP,1/1,DOUBLE

Figure 41. Schematic of one 400-gpm AquaSel module.

5.3 Implementation Scenarios

The product water from the AquaSel system, totaling 1,333 gpm or 2,154 acre-feet per year (AFY), would be returned to the desalters for production of potable water. Four alternatives were considered for how the AquaSel product water could be utilized. Although the AquaSel product water has a TDS similar to that of raw well water, returning it to the feed side of the desalters would not be a feasible alternative due to the high (~130 mg/L) concentration of silica in the product stream. This would increase the average silica concentration to the desalters and result in an overall decrease in the recovery.

5.3.1 Existing Desalter Streams and Flows

Although there are currently two desalters in operation, a third is being planned. Figure 42 presents a schematic of the arrangement of the existing desalters together with the future desalter (Perris II) and how their treated and bypass flow is combined, prior to final disinfection and distribution. Each desalter consists of two RO trains with a portion of bypass flow. In the case shown in Figure 42, about 250 gpm of raw water from the Menifee Desalter bypasses the RO trains and combines with the RO permeate downstream. Additionally, the Perris I Desalter would bypass about 700 gpm and the Perris II Desalter would bypass about 750 gpm. Product TDS from the desalters in this configuration is expected to be around 550 mg/L.

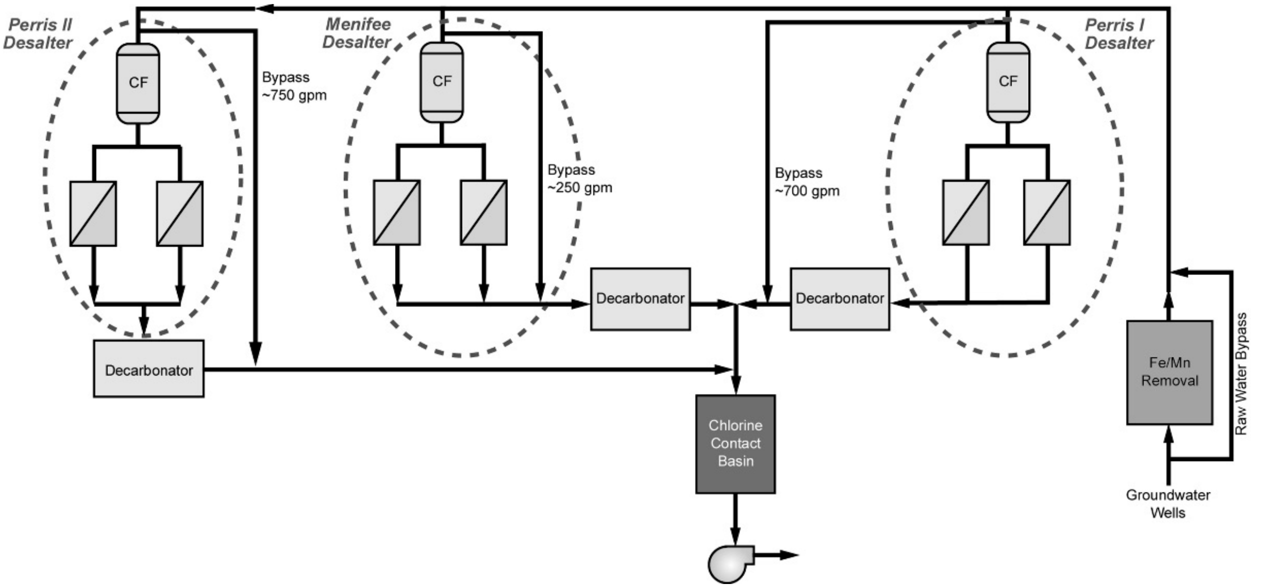


Figure 42. Schematic of existing desalter arrangement.

5.3.2 Alternative 1 – AquaSel Product Directly to Disinfection

For the first alternative the proposed AquaSel product stream would be combined with the product streams from the desalters just upstream of the chlorine contact basin. This configuration is illustrated in Figure 43.

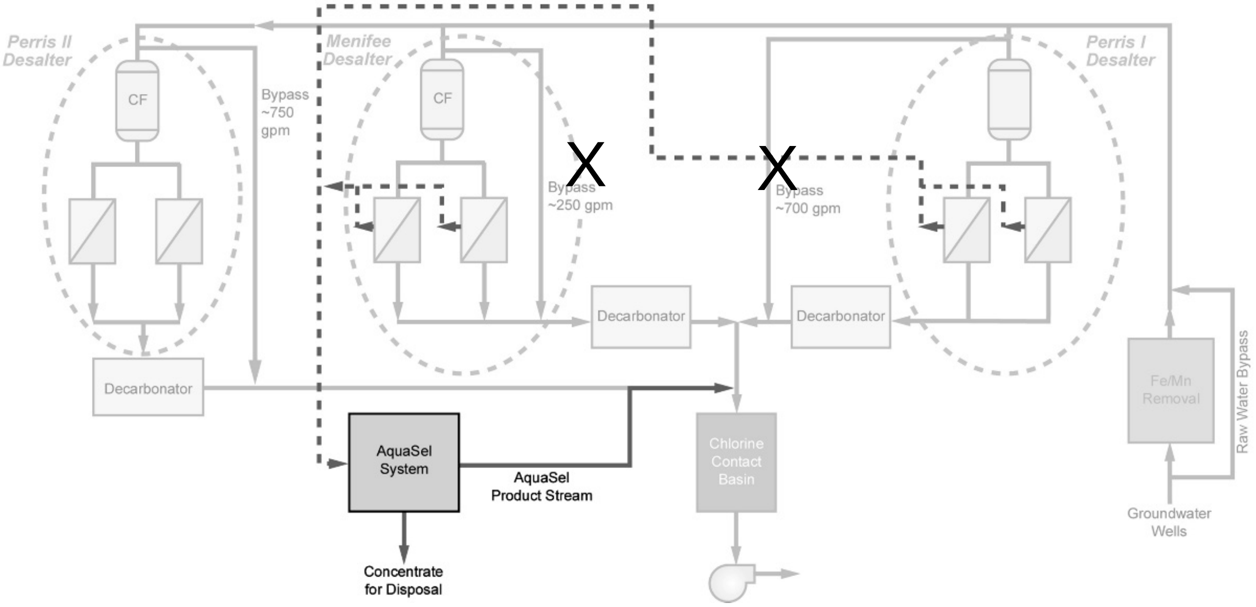


Figure 43. Alternative 1 — Blend AquaSel product with desalter product. Bypass around Menifee and Perris I off.

Because the AquaSel product water is expected to have a TDS concentration of around 1,600 mg/L, which is almost the same as the raw well water TDS, the bypass flow around the Menifee and Perris I Desalters would need to be turned off in order to maintain the finished water TDS concentration.

With the AquaSel product flow rate around 1,333 gpm and 950 gpm of bypass flow turned off, combining the AquaSel product flow with the permeate from the desalters would result in an overall increase in the product water production of about 383 gpm or 618 AFY, and an expected TDS in the combined finished potable water of around 670 mg/L.

This finished water TDS is slightly higher than the TDS of around 550 mg/L anticipated to be the combined TDS with all three desalters in operation.

5.3.3 Alternative 2 – AquaSel Product to Secondary EDR System

A second alternative would include adding an additional EDR treatment step downstream of the AquaSel system to polish the water further and produce a lower salinity stream that could be blended directly with the other product water. Figure 44 shows the process arrangement for this configuration. In this case, some water would be lost as concentrate blowdown from the secondary EDR process, but all bypass streams would remain, resulting in a higher overall product water yield. The resulting combined product TDS would be around 560 mg/L and the overall yield is expected to increase by 1,133 gpm or 1,828 AFY.

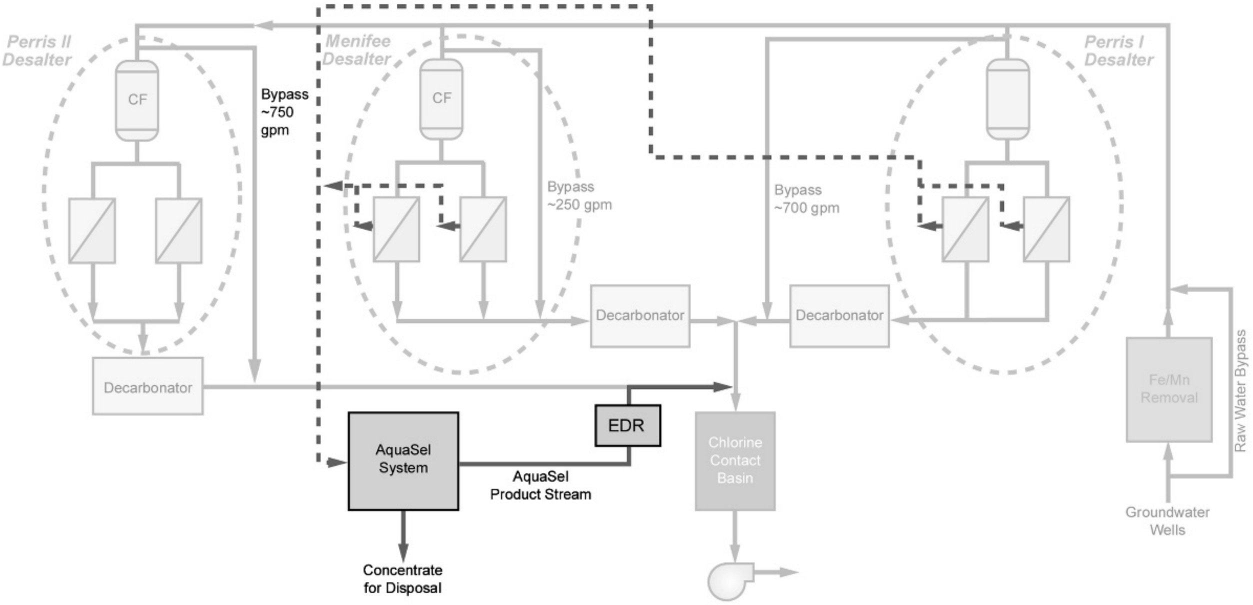


Figure 44. Alternative 2 — Secondary EDR treatment of AquaSel product to reduce TDS.

5.3.1 Alternative 3 – AquaSel Product Blending with Perris II RO Permeate

The third alternative is similar to Alternative 1, but in this case the AquaSel product water would be returned to blend with the permeate stream from the Perris II Desalter, in place of the raw water bypass. The configuration is shown in Figure 45. For this alternative, the resulting combined product TDS would be around 650 mg/L and the overall yield is expected to increase by only 583 gpm or 940 AFY. The yield is lower because the bypass around the Perris II Desalter would need to be turned off, to reduce the impacts on the product water TDS.

5.3.2 Alternative 4 – Replace Perris II RO with EDR and Blend AquaSel Product with Perris II EDR Feed

For Alternative 4, the Perris II treatment process would be changed from RO to EDR. There are two major process advantages of doing this. First, the EDR process is able to operate at a higher recovery rate on the raw well water (85 percent, compared to 75 percent for RO). Second, because EDR is not impacted by high silica concentrations in the feed stream, the AquaSel product could be recycled and used as an additional feed source for the Perris II EDR. This combination would therefore yield the highest product volume and also the lowest TDS product water. Estimates show that the product water TDS would be around 500 mg/L and the overall yield is expected to increase by 1,420 gpm or 2,291 AFY. The schematic configuration is shown in Figure 46.

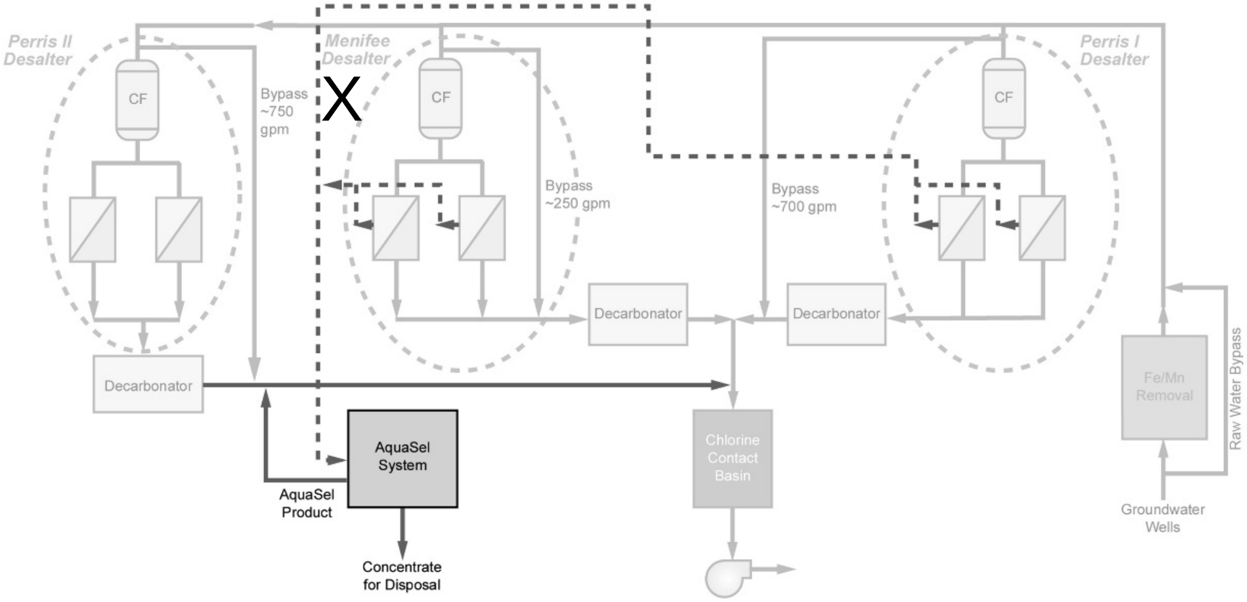


Figure 45. Alternative 3 — Blend AquaSel product with RO permeate from Perris II Desalter. Bypass stopped.

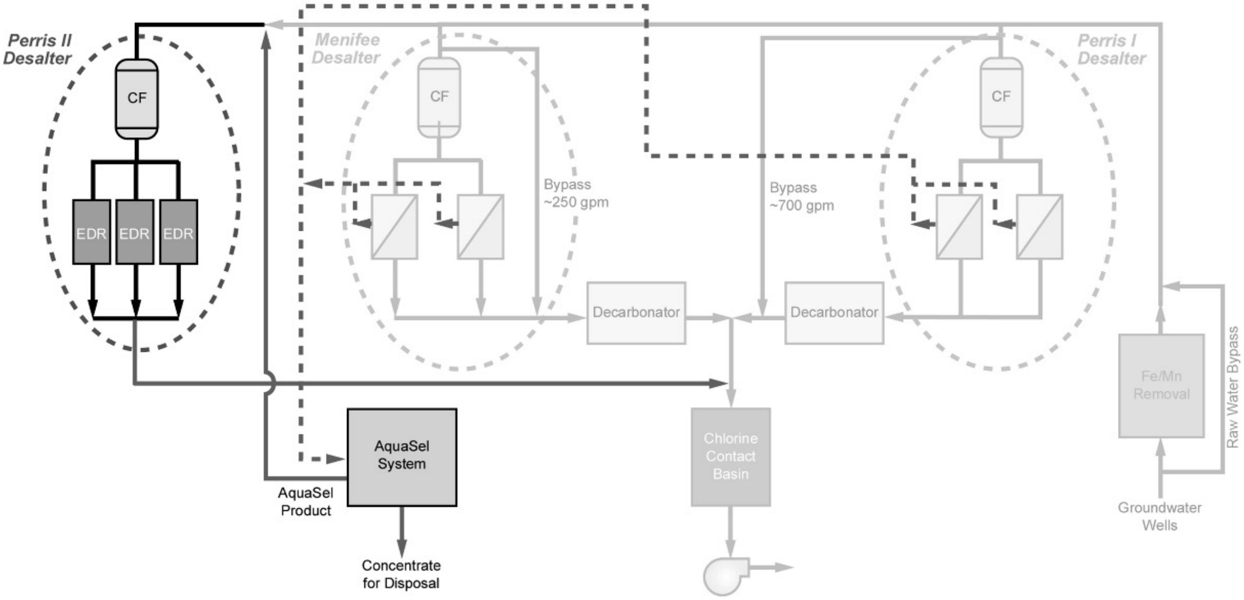


Figure 46. Alternative 4 — Return AquaSel product to front of Perris II EDR process.

5.4 Comparison of Alternatives

The above alternatives were compared on a cost basis to identify the most economically attractive alternative for implementing a full-scale AquaSel system for EMWD.

Capital costs were estimated to be between $40.1 million and $50.8 million depending on the alternative. O&M costs were estimated based on estimates for AquaSel operation, desalter operation, brine disposal costs, etc., as well as the annual cost of capital amortized at a 5-percent annual interest rate and a loan period of 30 years.

Table 28 summarizes the results. Also included in the table are total water production estimates for the alternatives, and the anticipated final product water TDS values.

Table 28. Comparison of Alternatives

Parameter	Alternative 1[1]	Alternative 2[2]	Alternative 3[3]	Alternative 4[4]
Total annual cost[5,6], $ M	19.5	20.8	19.5	20.9
Total water production, AFY	16,617	17,827	16,939	18,290
Anticipated combined finished water TDS, mg/L	~670	~560	~650	~500
Total estimated unit cost of water, $/AF	**1,383**	**1,371**	**1,356**	**1,346**
Total brine flow for final disposal, mgd	2.22	2.51	2.22	2.10

[1] Alternative 1 includes returning the AquaSel product stream to blend with the permeate from the existing Menifee and Perris I Desalters. The current practice of blending the RO permeate with bypass water would stop.

[2] Alternative 2 includes adding a secondary EDR process to reduce the TDS of the AquaSel product water to match that of the blended desalter product.

[3] Alternative 3 includes returning the AquaSel product stream to blend with the permeate from the Perris II Desalter. The bypass stream around the Perris II Desalter would be turned off.

[4] Alternative 4 includes replacing the Perris II RO process with EDR and returning the AquaSel product stream to the feed of the EDR.

[5] Based on a 5-percent interest rate and a 30-year loan period.

[6] Includes electrical power, chemicals, solids disposal, brine disposal, membrane replacement, equipment O&M, membrane and cartridge filter replacement, and additional labor costs.

As can be seen from the figures in Table 28, the relative costs of water from four alternative configurations that include the AquaSel process treating 2.4 mgd of RO concentrate are similar; in the range of $1,346 to $1,383 per acre-foot (AF). The lowest cost alternative is Alternative 4, which includes EDR as the Perris II Desalter process. This is because that system would result in a higher overall recovery and, therefore, the highest total water production and the lowest volume of brine requiring final disposal.

6.0 SUMMARY AND CONCLUSIONS

6.1 Summary

Over a 9-month period, between November 2014 and July 2015, EMWD tested GE's AquaSel technology on RO concentrate from the Menifee groundwater desalter. The test work was undertaken in three separate phases. Phases 1 and 2 were focused on troubleshooting different aspects of the process and optimizing the performance. Phase 3 focused on achieving up to 1,000 hours of operation at one set of operating conditions in the AquaSel and with the Menifee Desalter operating at 75 percent recovery. A total of 962 operating hours were achieved during Phase 3, and throughout that period the AquaSel system operated at 80 percent recovery, resulting in an overall recovery of 95 percent for the system. The EDR system performance was stable and produced a product stream with a quality similar to that of the raw groundwater that feeds the desalters. The precipitator performance, however, was not consistently stable. This led to difficulties in obtaining a measured mass balance over the system and resulted in a buildup of solids in the return pipework and filters downstream of the precipitator.

There are several ways in which the AquaSel system can be incorporated into EMWD's existing and future desalting configuration. The relative costs for each alternative approach are similar, but some configurations yield a greater total water production, lower final product water TDS, and lower volumes of brine for disposal than others.

6.2 Progress with Respect to Project Goals

The overall purpose of this pilot study was to determine the feasibility of secondary treatment of concentrate from the Menifee and Perris I Desalters to reduce the volume of brine for final disposal and increase the overall recovery of potable water to around 95 percent or greater. The pilot study has determined that it is feasible to treat the existing RO concentrate and achieve an overall recovery of 95 percent.

The progress toward meeting specific pilot plant objectives was assessed based on the performance during Phase 3 of the testing period, and these assessments are presented below.

6.2.1 Obtain Consistent, Steady-State Operation

As mentioned, the EDR system demonstrated consistent and steady-state operation during Phase 3 testing. The objective was to operate at one set of conditions for about 1,000 hours. This goal was met, and a total operating time of 962 hours was achieved during Phase 3.

During Phase 3, 16 sets of analytical data were obtained to monitor the water quality in various process streams. Specifically, the EDR system pressures, EDR system voltages and amperages, and the feed, product, and concentrate conductivities were monitored continuously.

6.2.2 Solids Production Rate

The performance of the precipitator tank was not as stable as that of the EDR system. Consequently, a consistent solids production rate was not obtained. Measurement of parameters in various process streams facilitated the estimation of solids production.

The chemical makeup of the solids produced by the process was estimated to be most likely a mixture of gypsum and calcium carbonate; however, this conclusion was not confirmed by analytical testing.

6.2.3 Anticipated Long-Term Operating Costs

The consumption rates for all chemicals used in the pilot plant were monitored and recorded. These values, together with estimates of the electrical power costs, membrane cleaning frequency and costs, solids production rate and disposal cost, and membrane and other EDR stack component replacement costs were extrapolated to estimate the cost of a full-scale system. Annual costs for a plant to treat 2.4 mgd were estimated to be $3.1 million.

6.3 Overall Conclusions

Based on pilot testing of the GE AquaSel process presented in this report, the following conclusions were reached:

1. EMWD was able to demonstrate successful operation of the 8-gpm AquaSel pilot system while operating on RO concentrate from the Menifee Desalter.
2. During 962 hours of operation at one set of operating conditions during Phase 3 of the testing, the AquaSel system was able to operate at 80 percent recovery. When combined with the 75 percent recovery of the Menifee Desalter, this resulted in an overall recovery of 95 percent.
3. Performance of the EDR stack during Phase 3 was stable, and it produced water similar in quality to that of the raw groundwater that feeds the desalters. The average TDS of the product water was 1,618 mg/L, with an average silica concentration of 131 mg/L. The average salt rejection of the EDR stack was in the range of 70 to 75 percent.

4. Performance of the Precipitator tank was not consistent. This led to difficulties in obtaining a true mass balance across the system, excess solids in the overflow stream returning to the EDR stack, and frequent replacement of cartridge filters. This is an area of the process that requires some modifications if additional test work is planned.
5. The construction cost (2015) for an AquaSel system to treat 2.4 mgd of brine from the Menifee and Perris I Desalters is estimated to be $31.4 million. The total project cost is expected to be around $40 million.
6. Annual operating costs are estimated to total about $3.1 million, with over 55 percent of that cost associated with electrical power and chemicals. The annual operating costs include almost $500,000 for disposal of the final concentrate stream produced by the AquaSel system. Before disposal, this stream would be diluted with excess recycled water to prevent scale formation in EMWD's brine pipeline. Another $416,000 would be required for the annual disposal of solids produced by the system.
7. The AquaSel product water could be either:
    a. Combined with the desalter permeate streams to replace the current use of bypass flow (Alternative 1);
    b. Treated by a secondary EDR system to reduce the TDS so that it can become part of the overall product stream without sacrificing the bypass flows (Alternative 2);
    c. Combined with the permeate flow from the Perris II Desalter to replace the use of bypass flow (Alternative 3); or
    d. Combined with the feed stream to an EDR based Perris II Desalter, which would be able to deal with the higher silica concentration in the AquaSel product stream.

   A cost comparison showed that all four alternatives resulted in a similar overall cost of water from the desalter system. Alternative 4 had the lowest cost because it achieved the highest total water production and the lowest flow of brine for final disposal. The average unit cost of water for the alternatives ranged from $1,346/AF for Alternative 4 to $1,383/AF (for Alternative 1.

Based on the promising performance of the AquaSel system during the pilot testing, EMWD is considering implementation of a 100-gpm AquaSel demonstration project to confirm the results of the pilot testing and focus on testing improvements to the precipitation system and downstream solids removal process.

7.0 REFERENCE LIST

Bureau of Reclamation, April 2008. *Evaluation and Selection of Available Processes for a Zero-Liquid Discharge System for the Perris, California, Ground Water Basin*, Desalination and Water Purification Research and Development Program Report No. 149, Denver, Colorado.

Santa Ana Watershed Project Authority, May 2010. *Santa Ana Watershed Salinity Management Program – Phase 2 SARI Planning Technical Memorandum*, Report by CDM, Carollo and Wildermuth, Riverside, California.

APPENDIX A.1: EMWD PHASE 3 EDR CONDUCTIVITY CHARTS

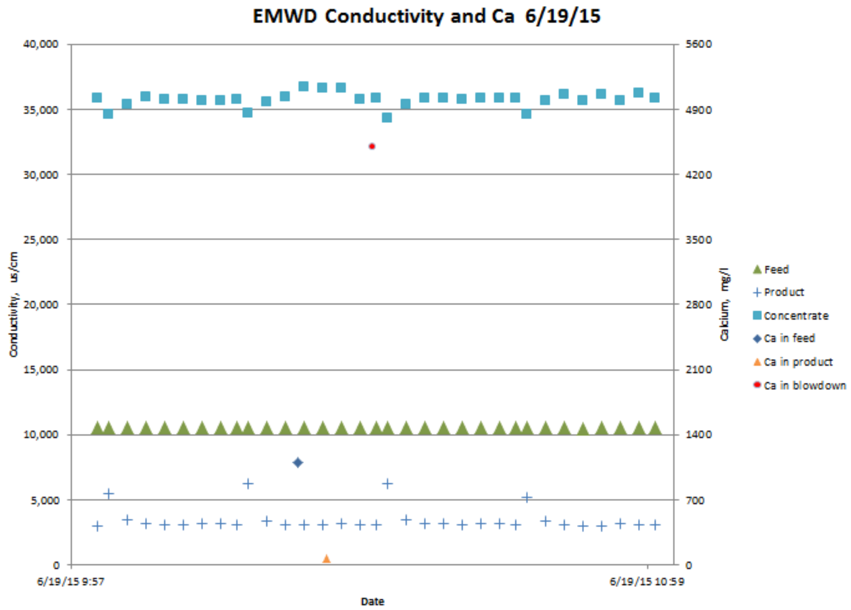


Figure A1. Conductivity and Ca 6/19/2015.

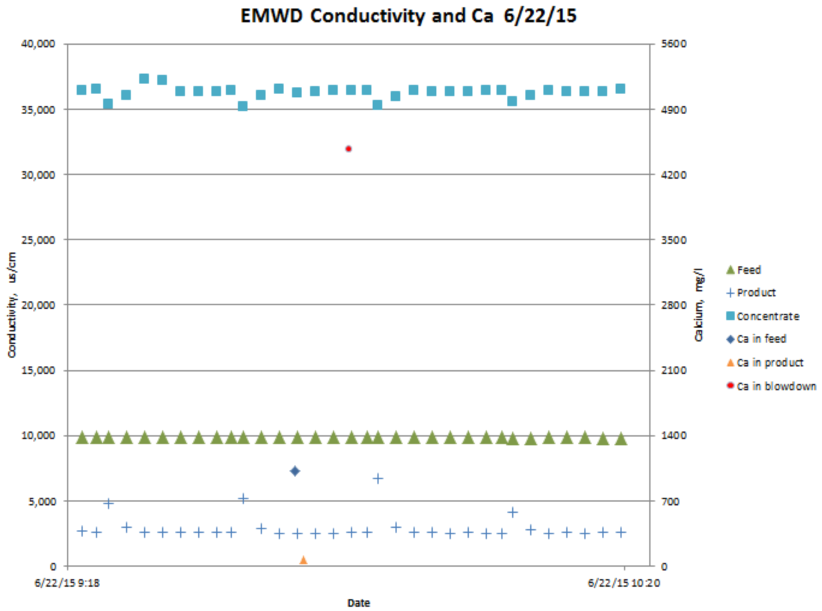


Figure A2. Conductivity and Ca 6/22/2015.

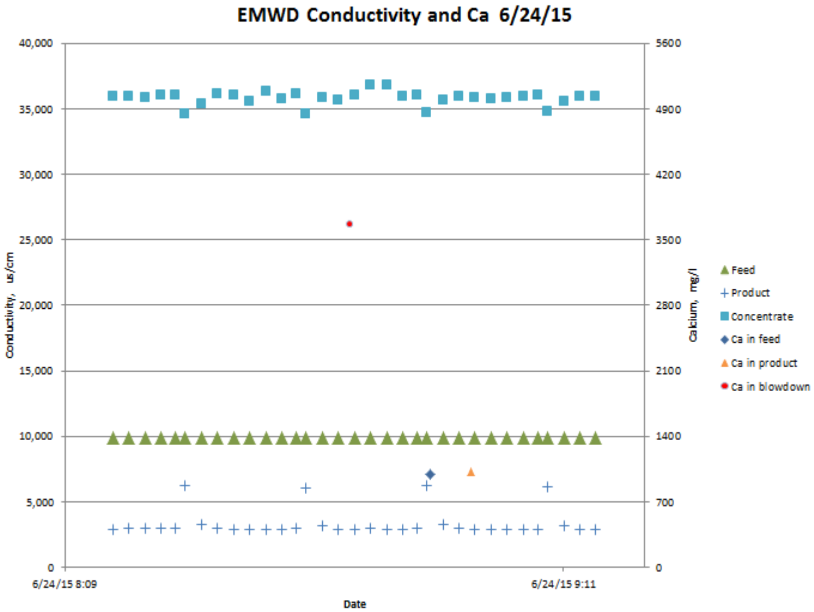


Figure A3. Conductivity and Ca 6/24/2015.

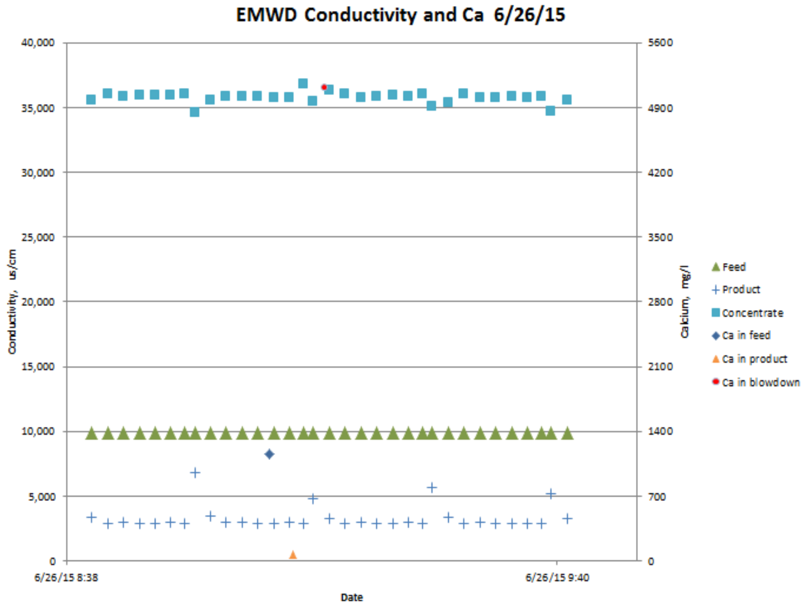


Figure A4. Conductivity and Ca 6/26/2015.

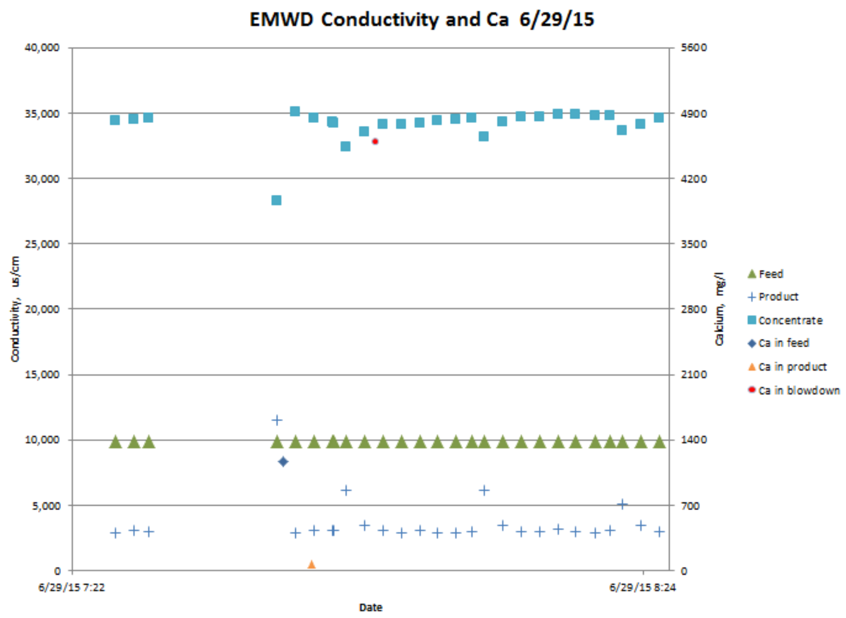


Figure A5. Conductivity and Ca 6/29/2015.

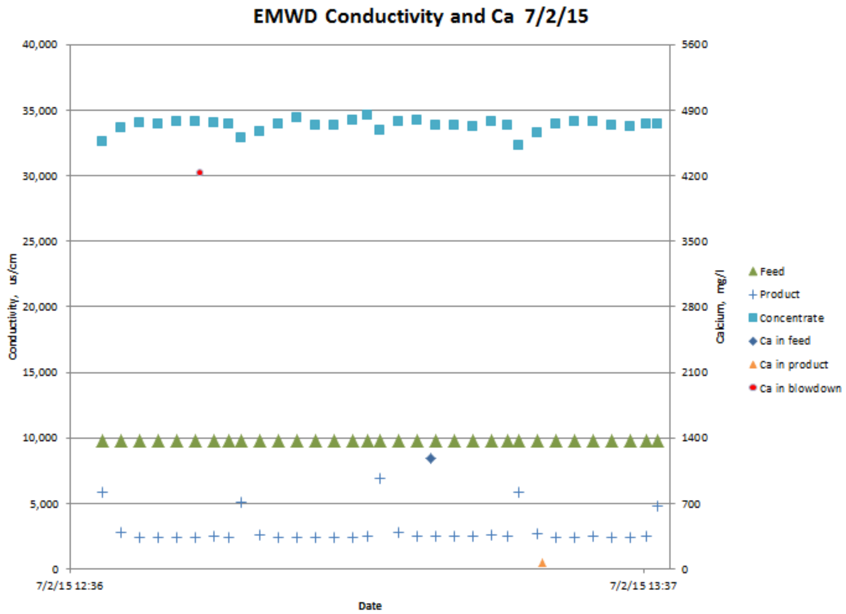


Figure A6. Conductivity and Ca 7/2/2015.

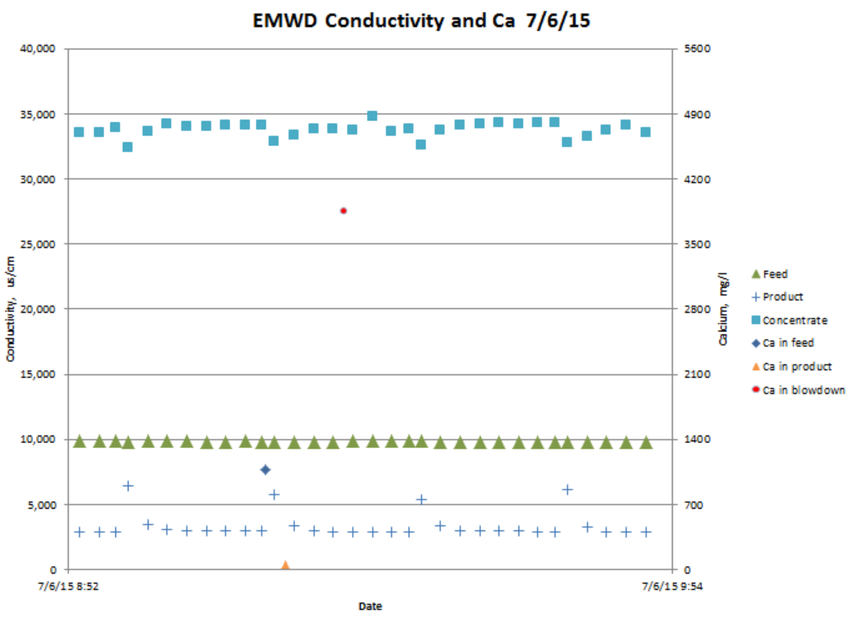
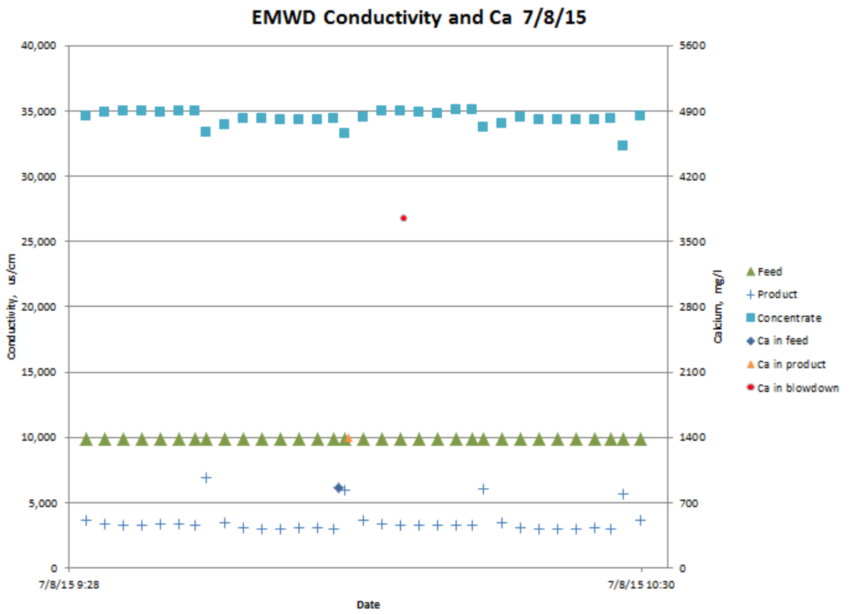


Figure A7. Conductivity and Ca 7/6/2015.



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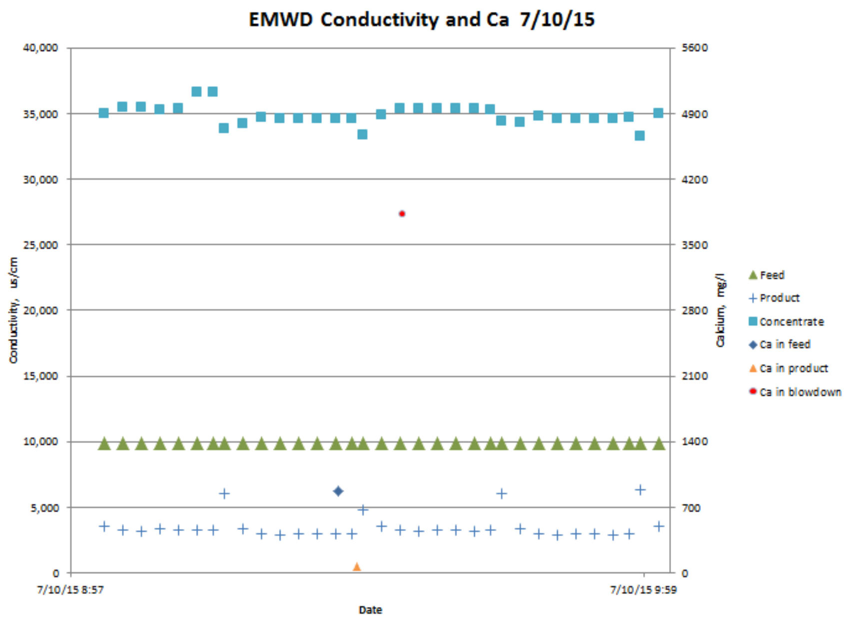

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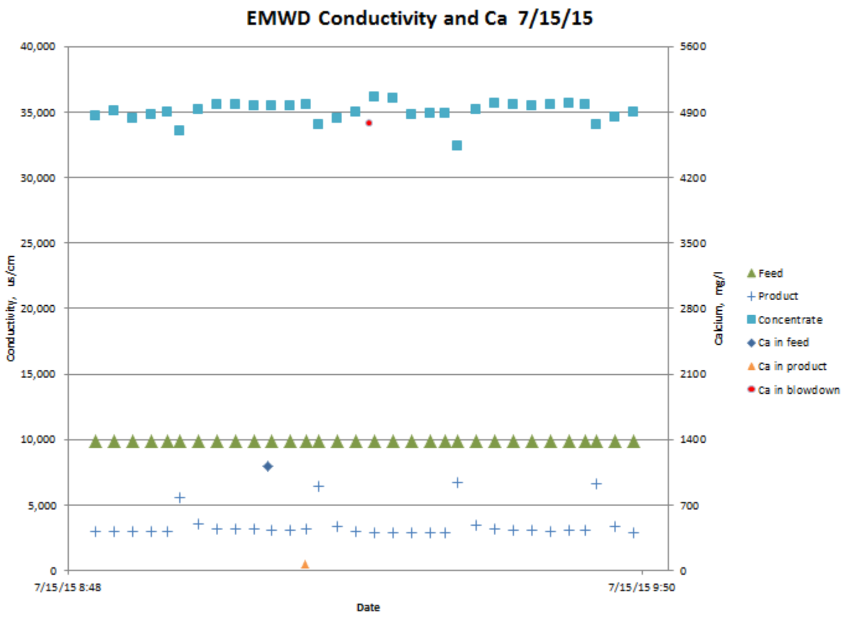


Figure A11. Conductivity and Ca 7/15/2015.

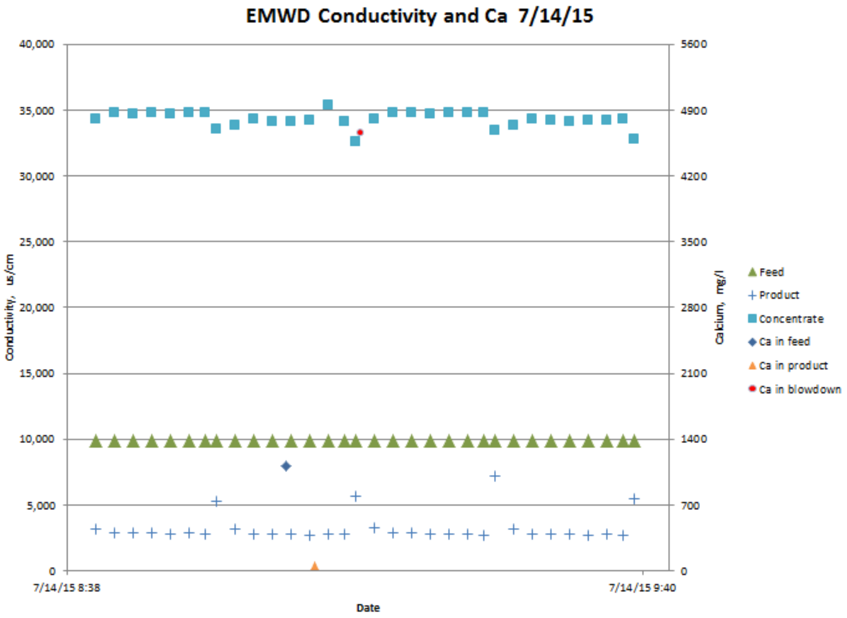
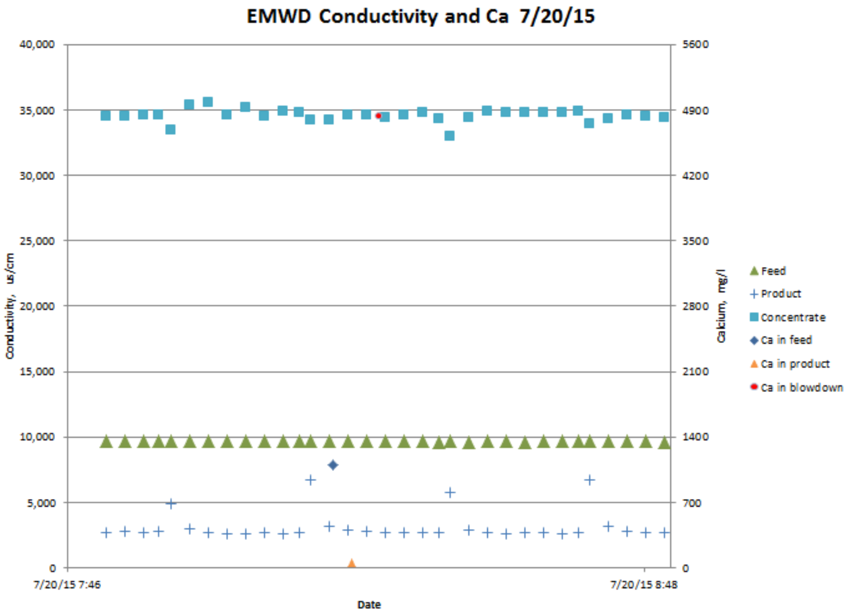


Figure A10. Conductivity and Ca 7/14/2015.



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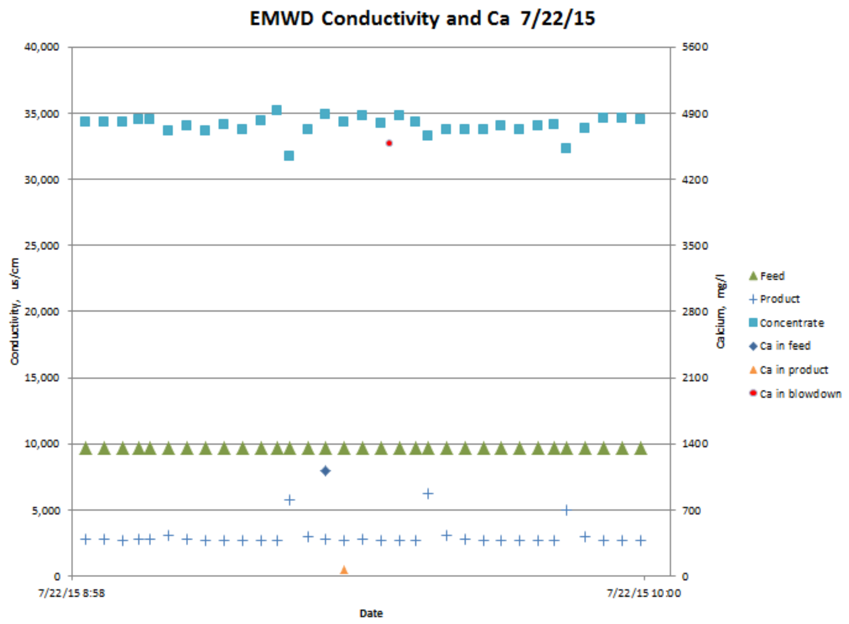

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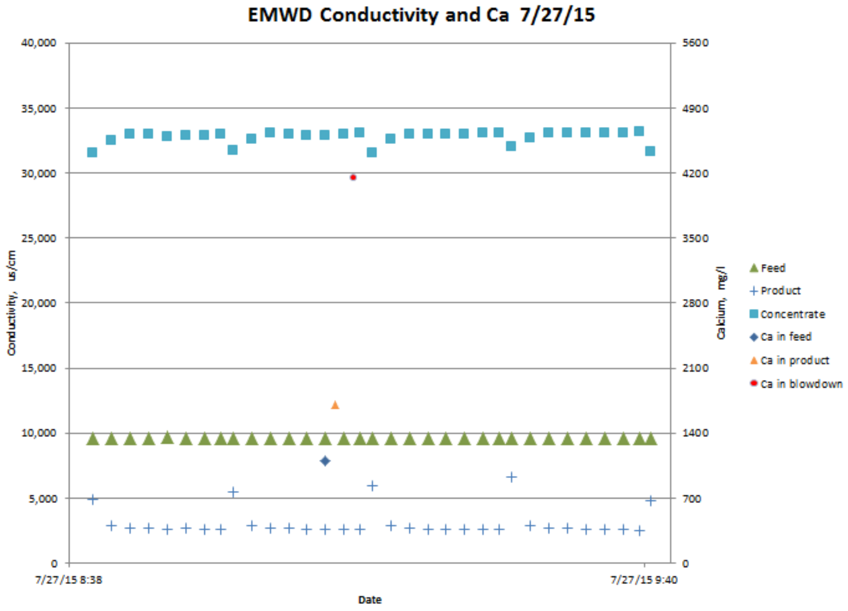

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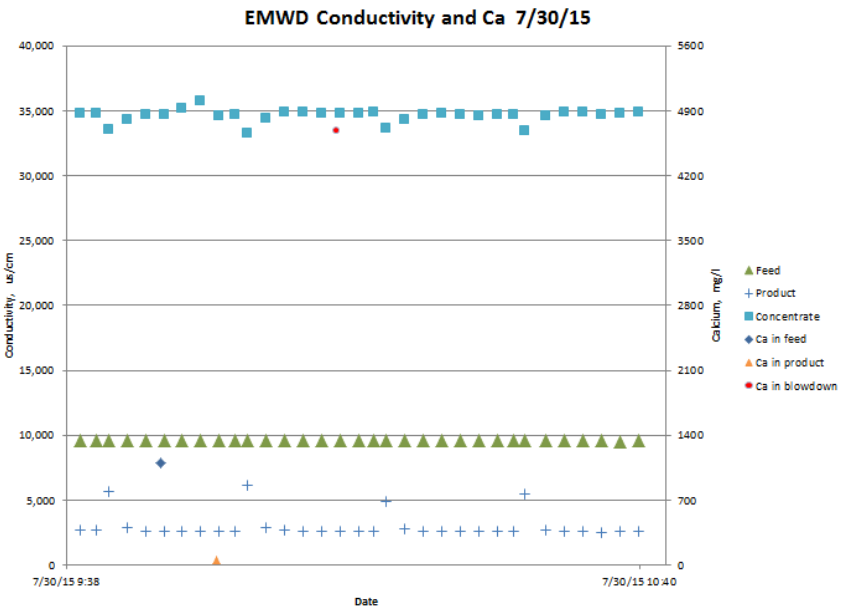

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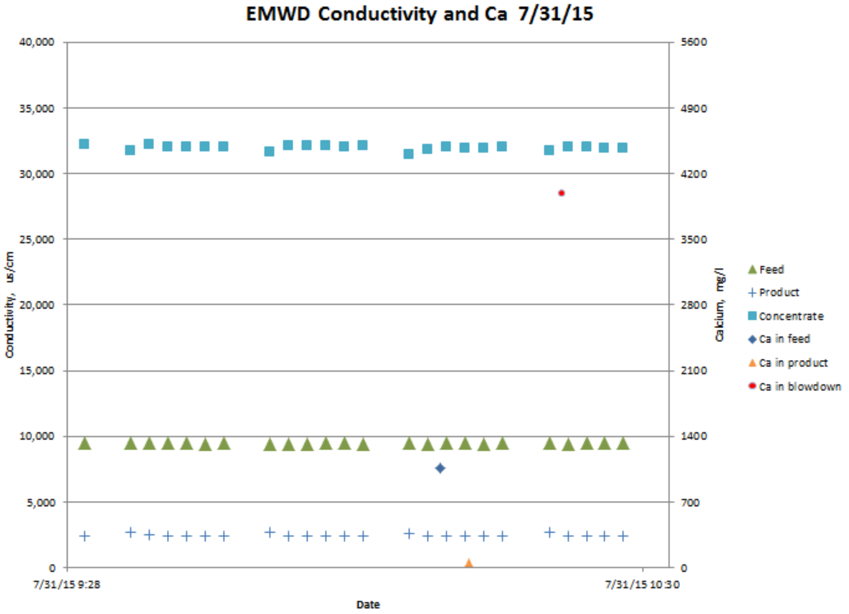


Figure A16. Conductivity and Ca 7/31/2015.

APPENDIX A.2: EMWD PHASE 3 EDR FLOW CHARTS

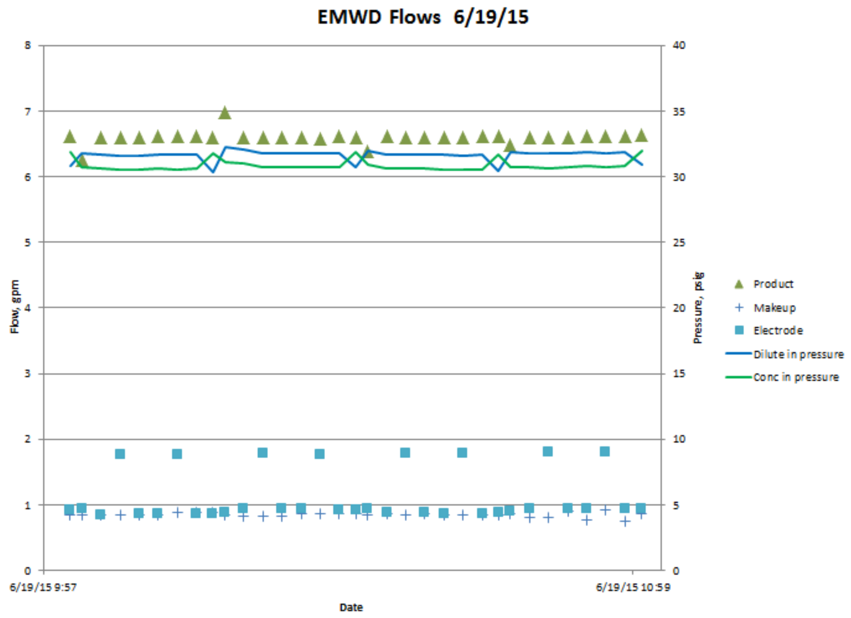


Figure A17. Flow snapshot 6/19/2015.

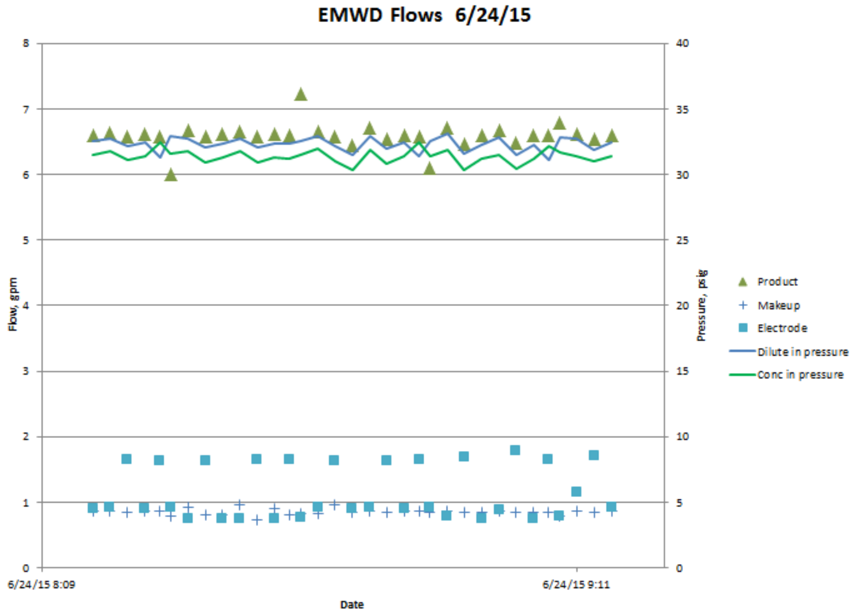


Figure A19. Flow snapshot 6/24/2015.

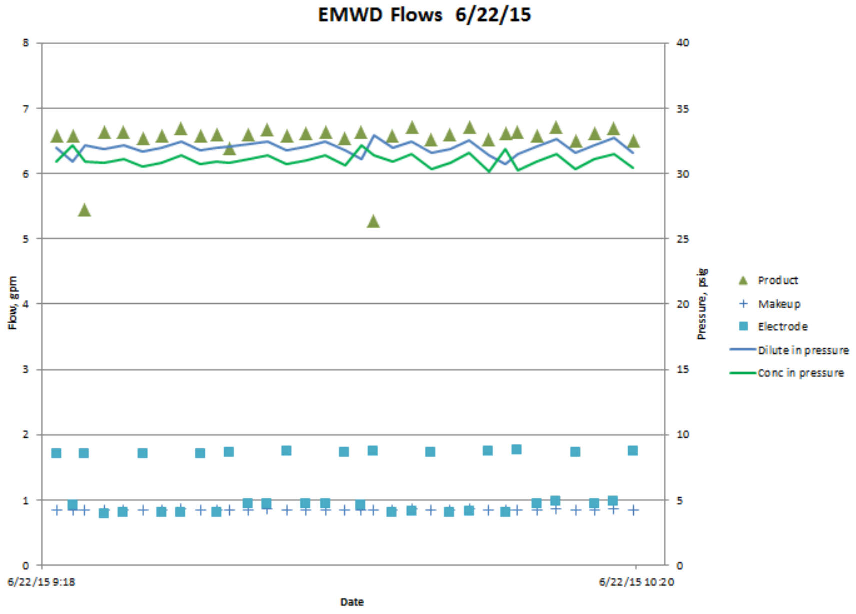


Figure A18. Flow snapshot 6/22/2015.

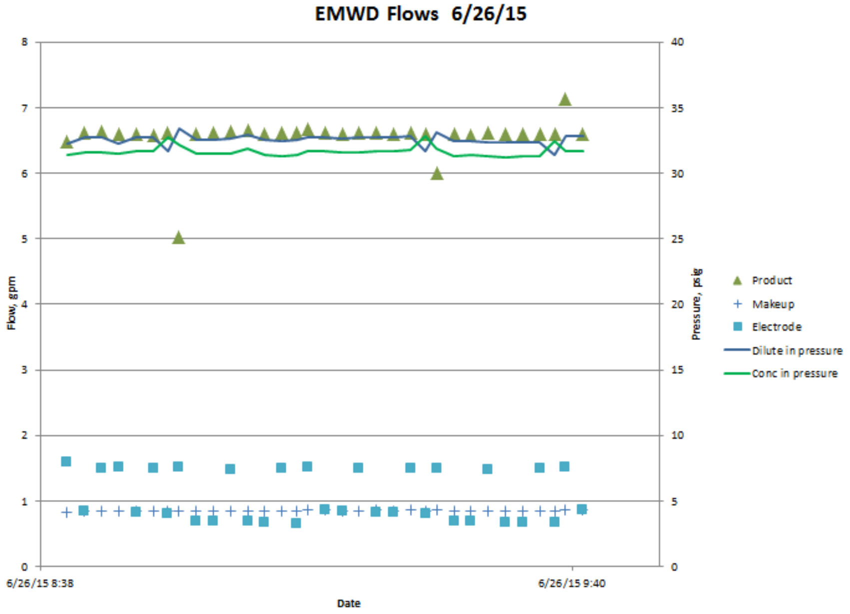
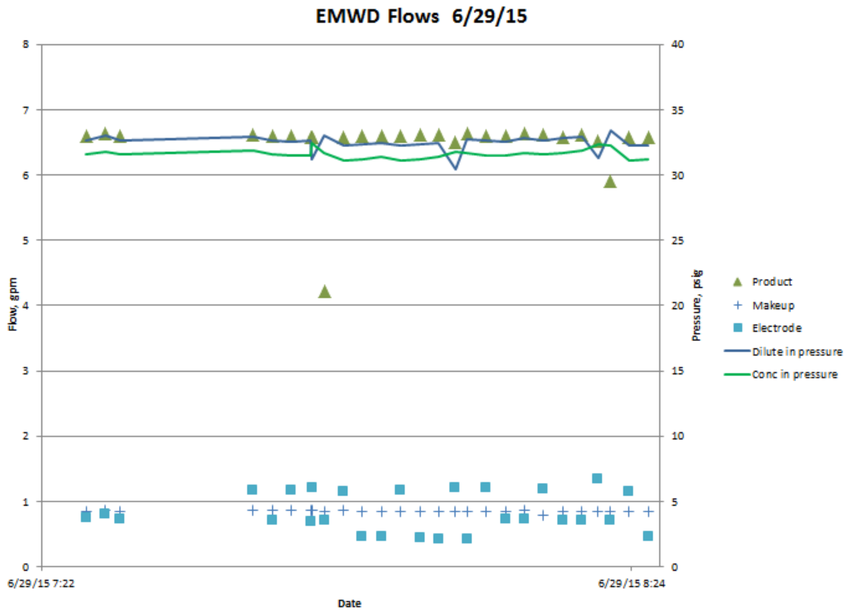


Figure A20. Flow snapshot 6/26/2015.



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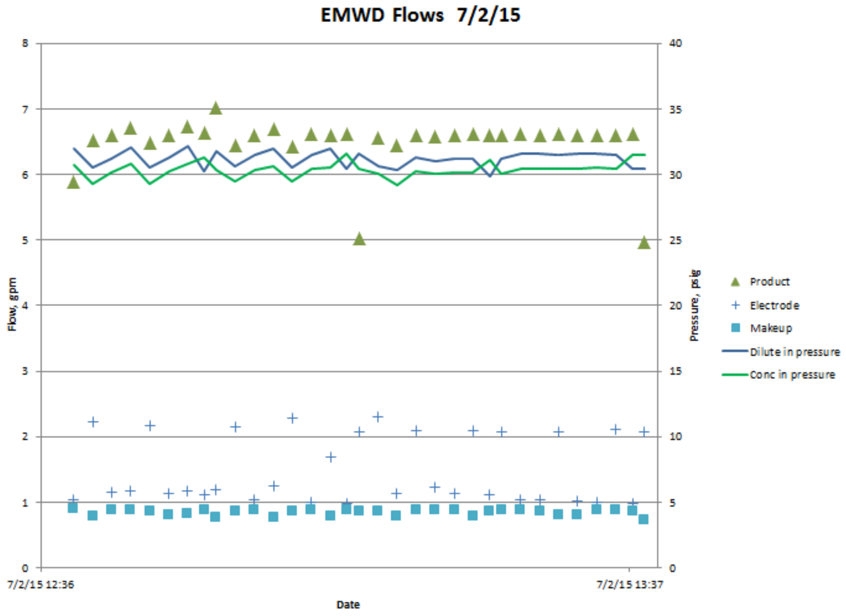
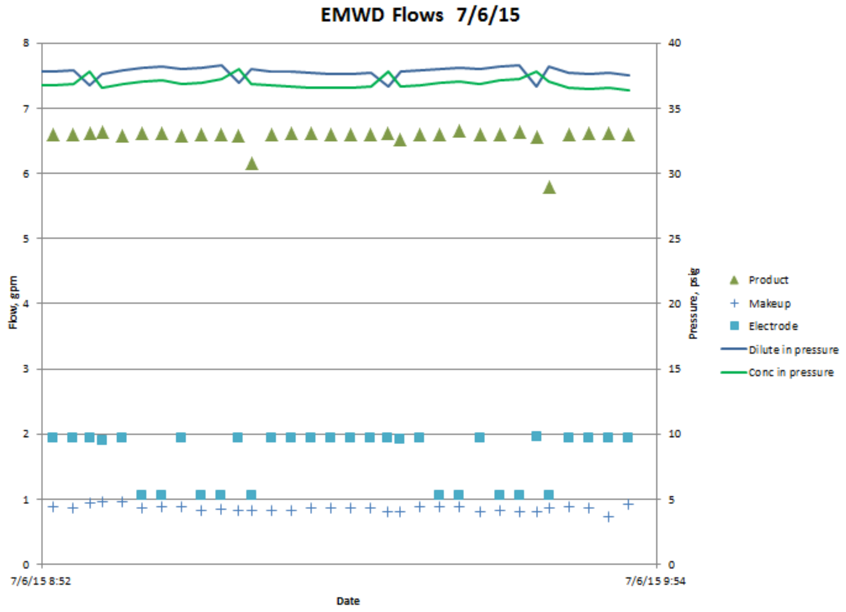


Figure A22. Flow snapshot 7/2/2015.



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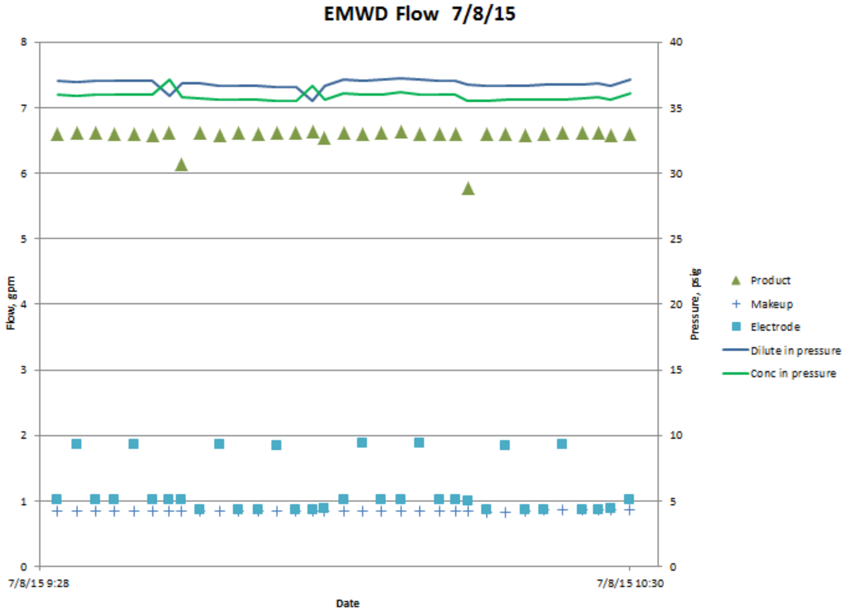

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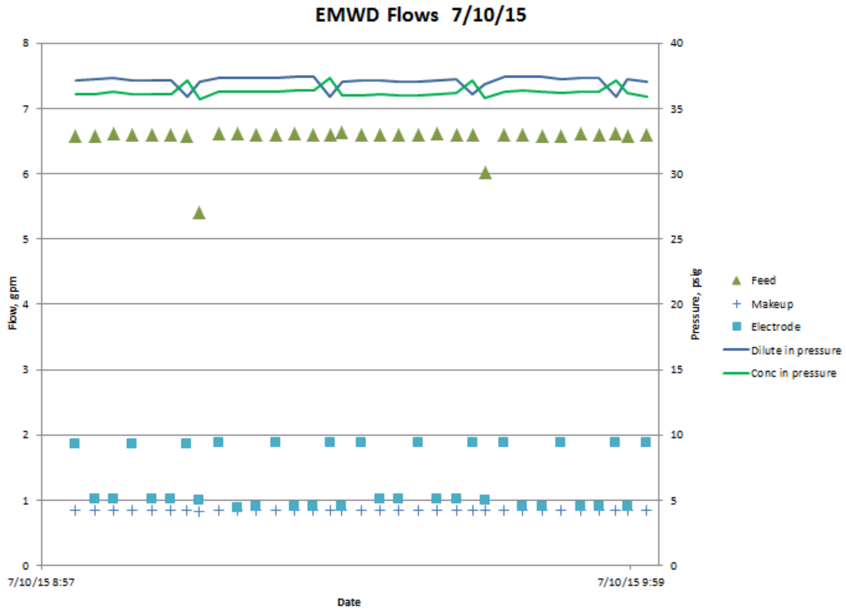

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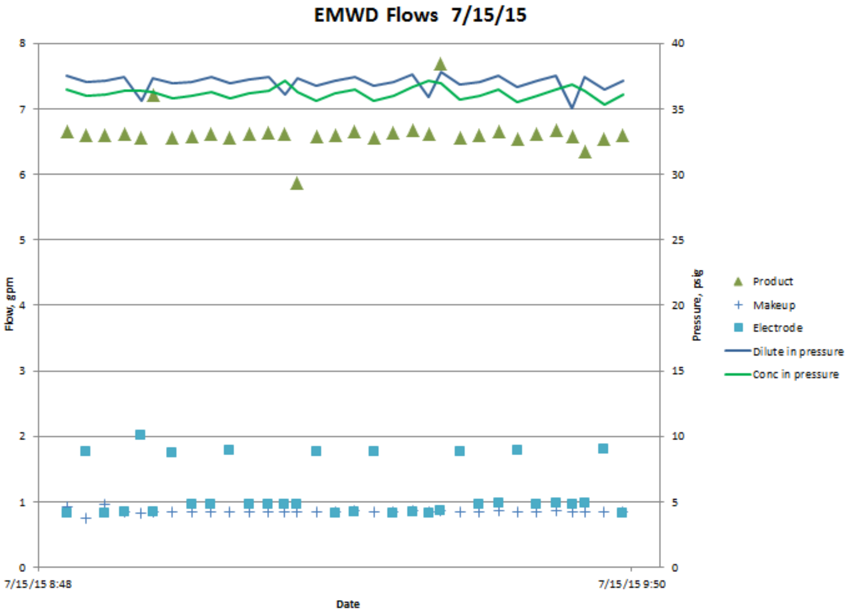


Figure A27. Flow snapshot 7/15/2015.

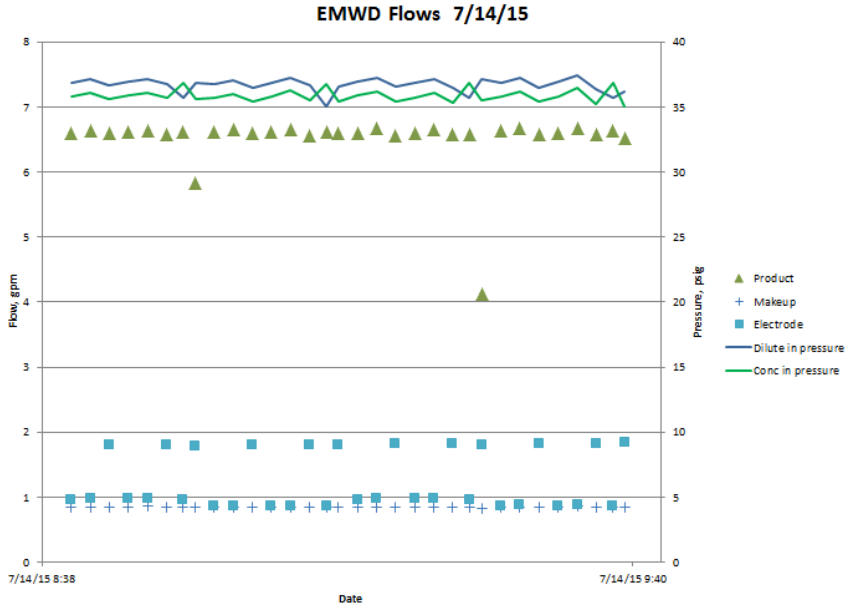


Figure A26. Flow snapshot 7/14/2015.

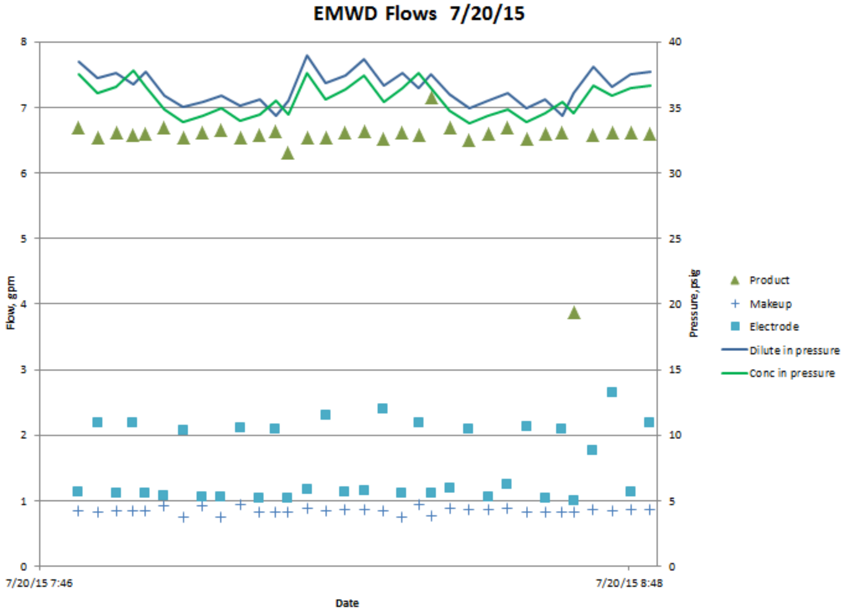
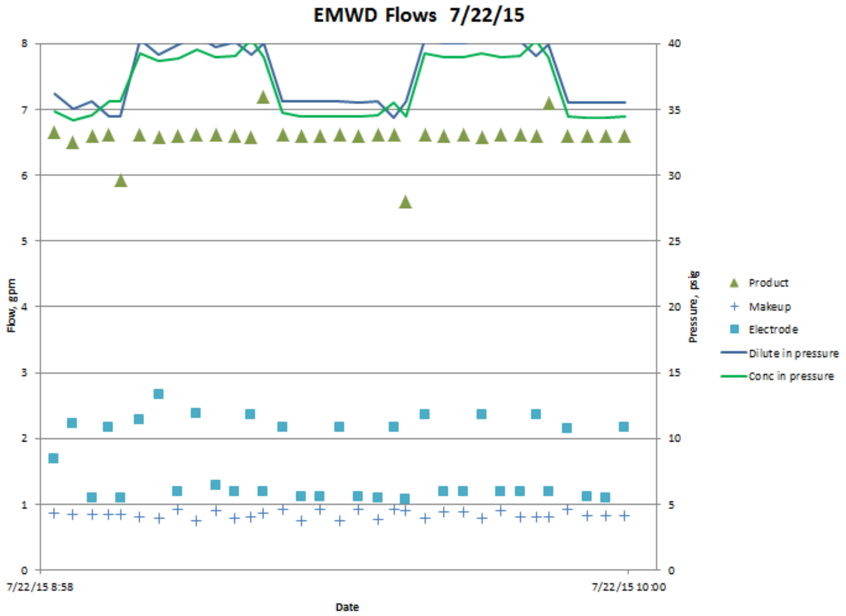


Figure A28. Flow snapshot 7/20/2015.



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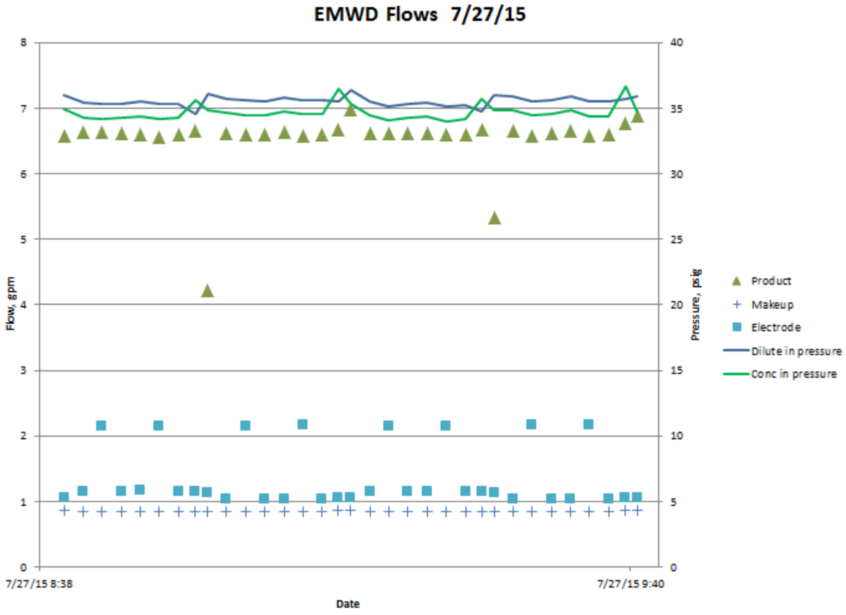
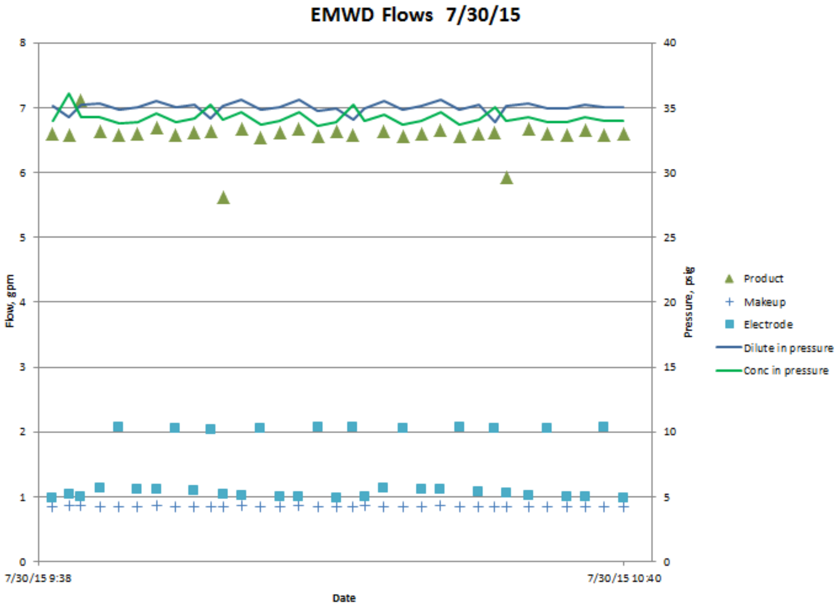


Figure A30. Flow snapshot 7/27/2015.



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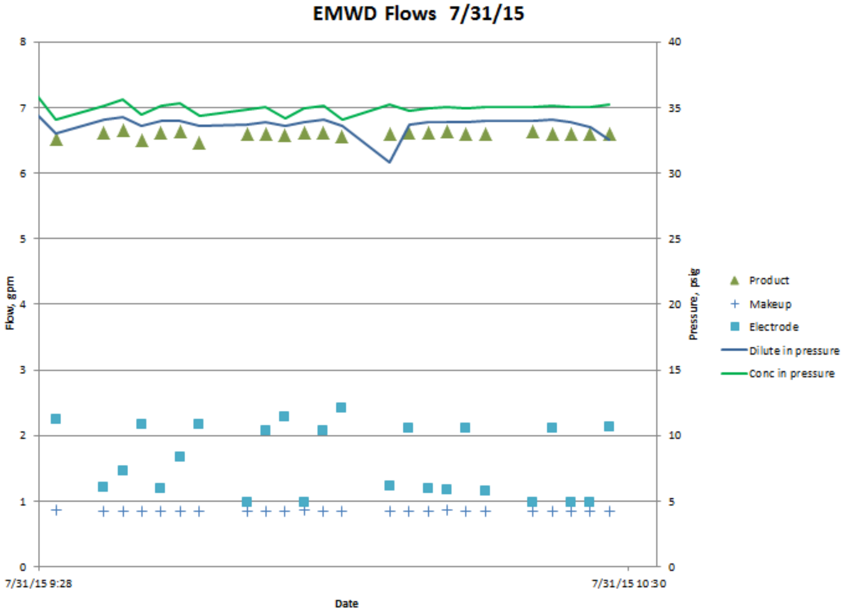

Figure A32. Flow snapshot 7/31/2015.

APPENDIX A.3: EMWD PHASE 3 EDR VOLTAGE AND CURRENT CHARTS

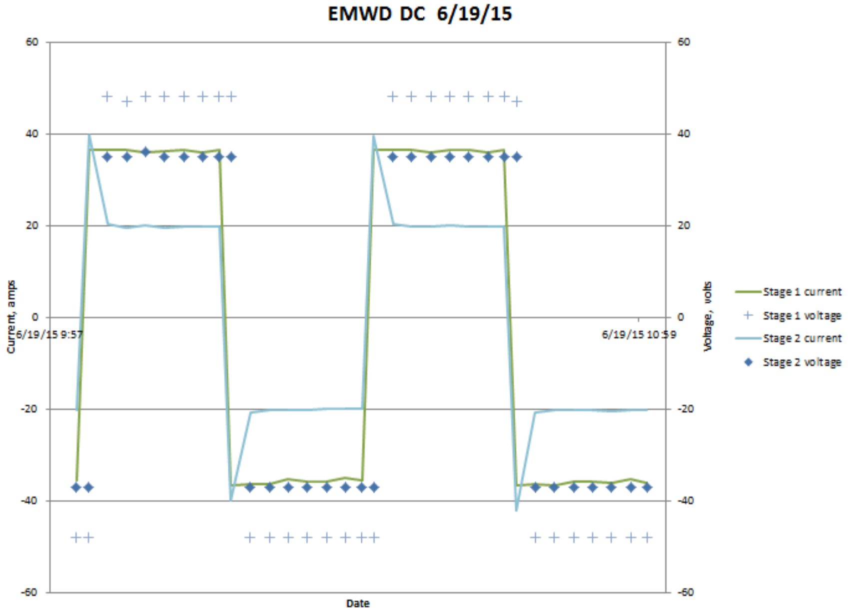

Figure A33. Voltage and amperage 6/19/2015.

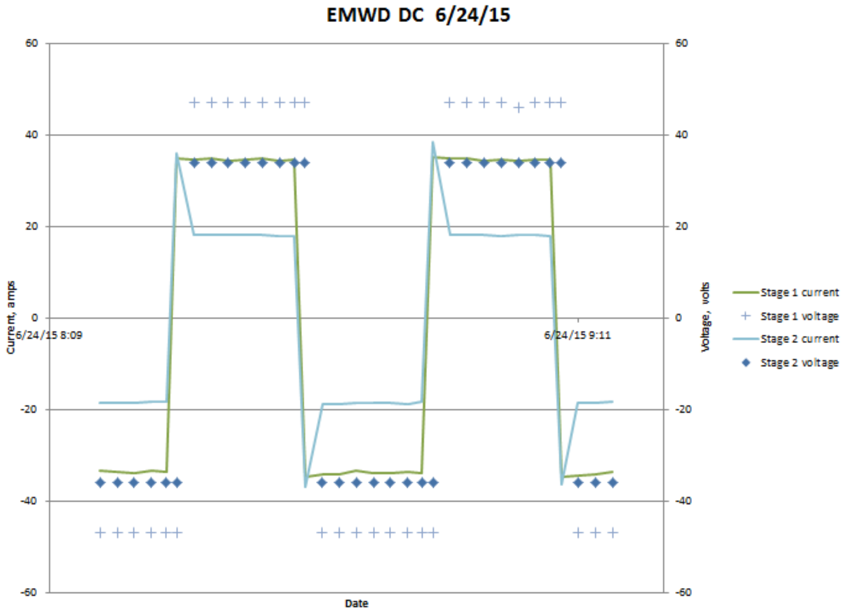

Figure A35. Voltage and amperage 6/24/2015.

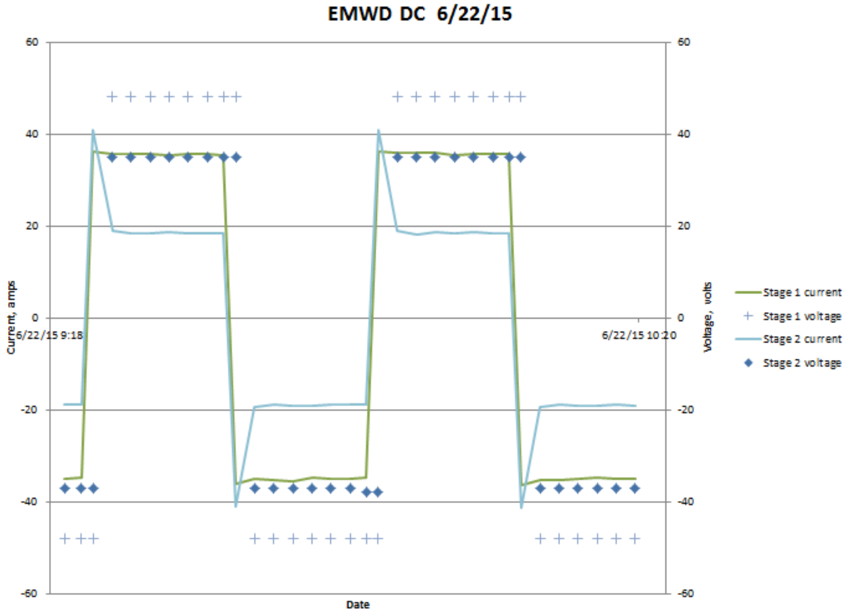

Figure A34. Voltage and amperage 6/22/2015.

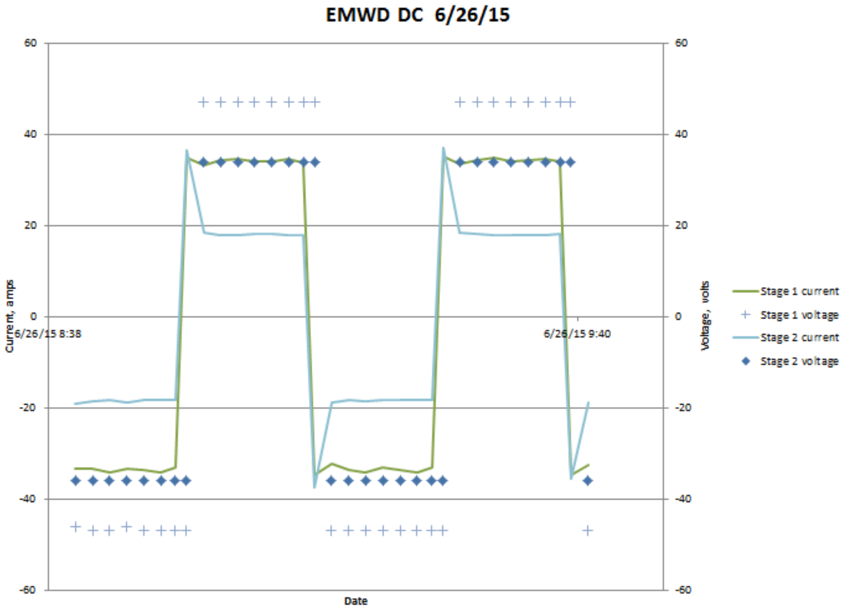

Figure A36. Voltage and amperage 6/26/2015.

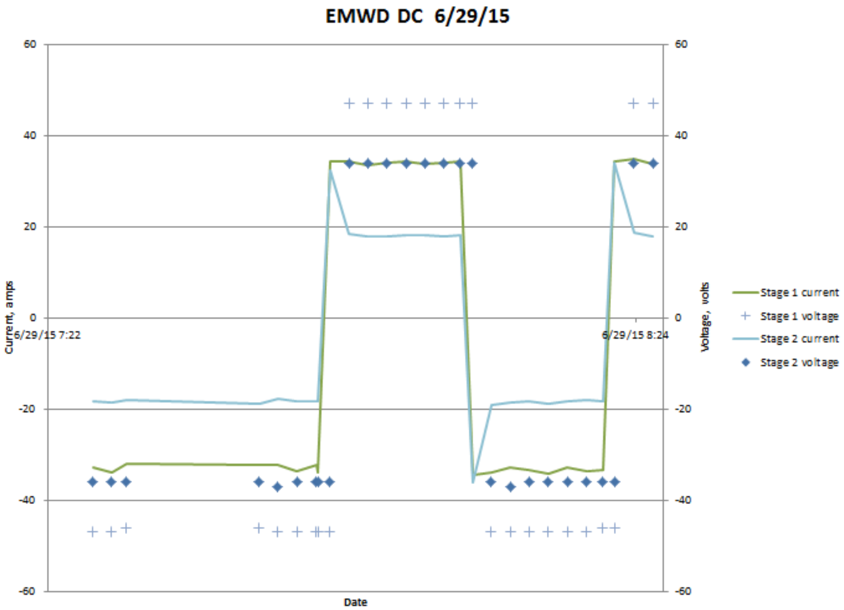

Figure A37. Voltage and amperage 6/29/2015.

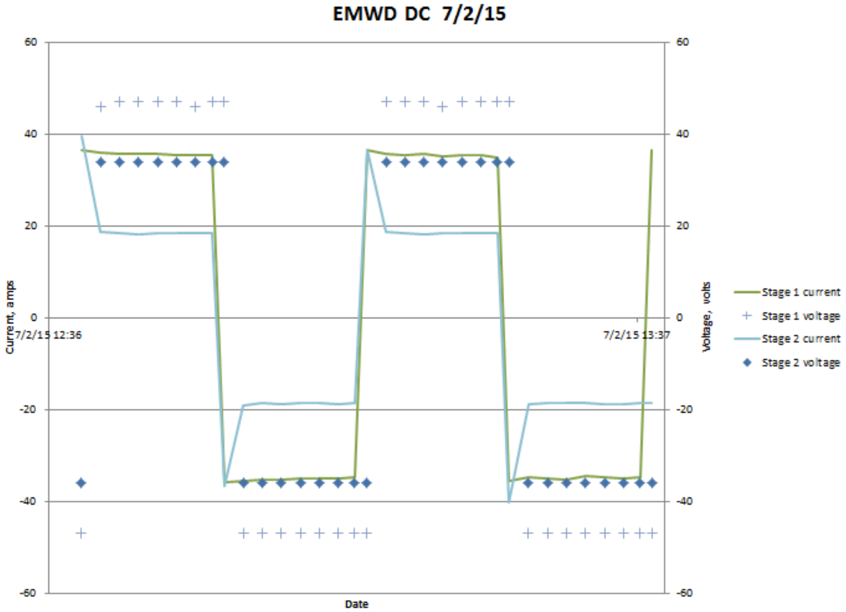

Figure A38. Voltage and amperage 7/2/2015.

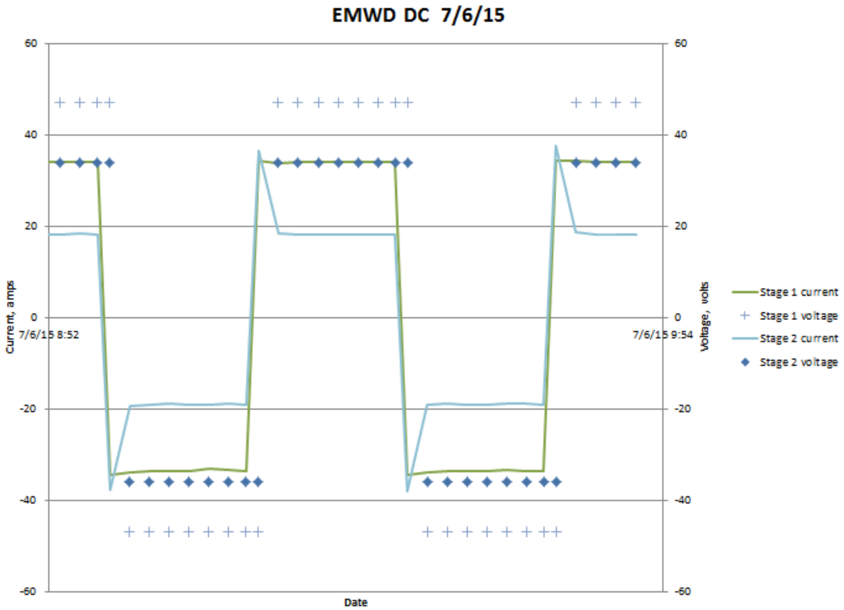

Figure A39. Voltage and amperage 7/6/2015.

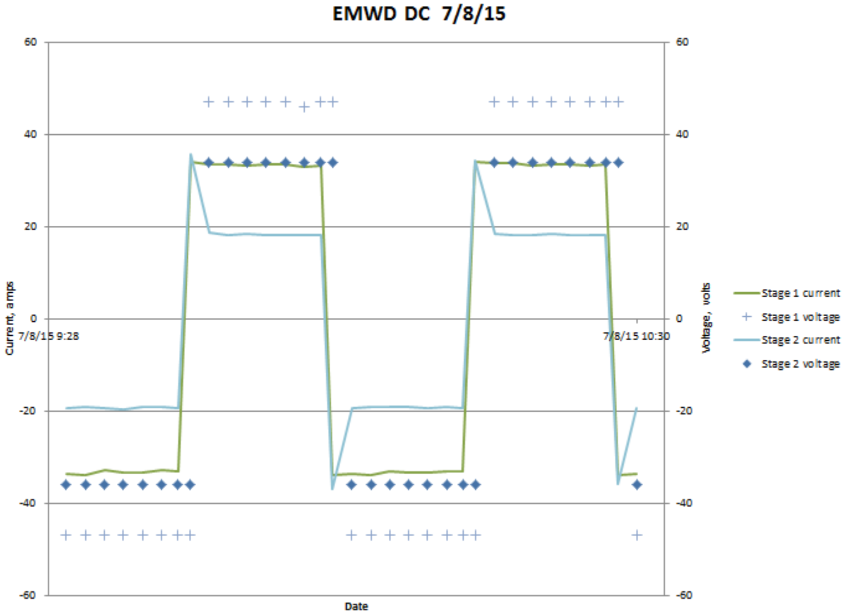

Figure A40. Voltage and amperage 7/8/2015.

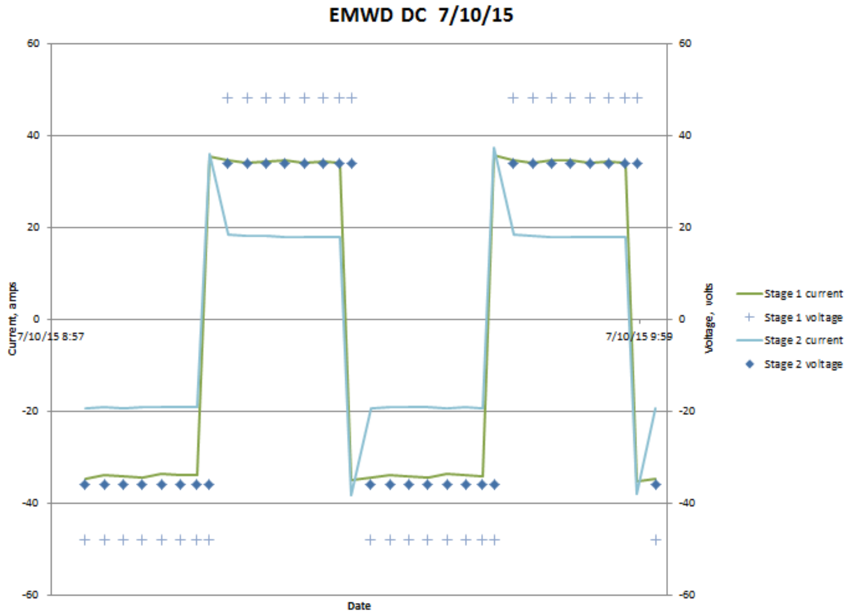

Figure A41. Voltage and amperage 7/10/2015.

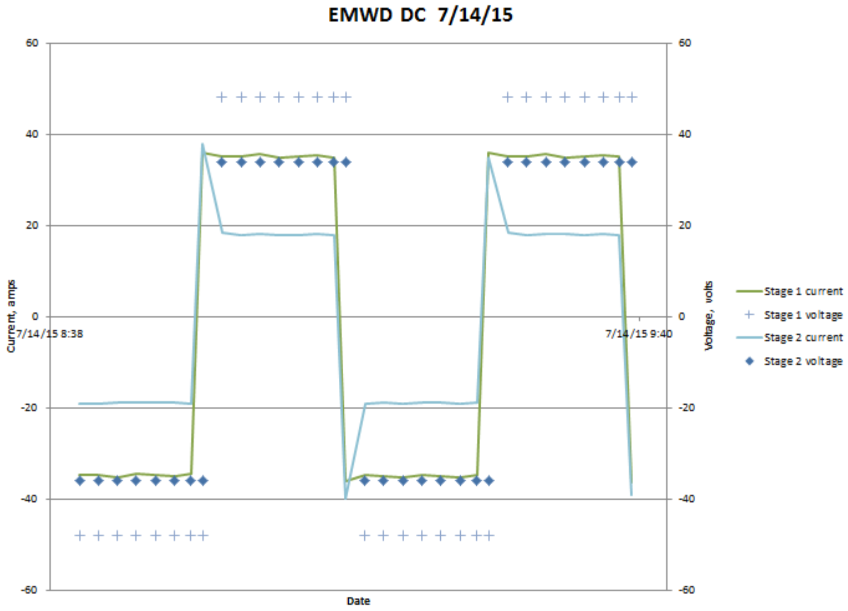

Figure A42. Voltage and amperage 7/14/2015.

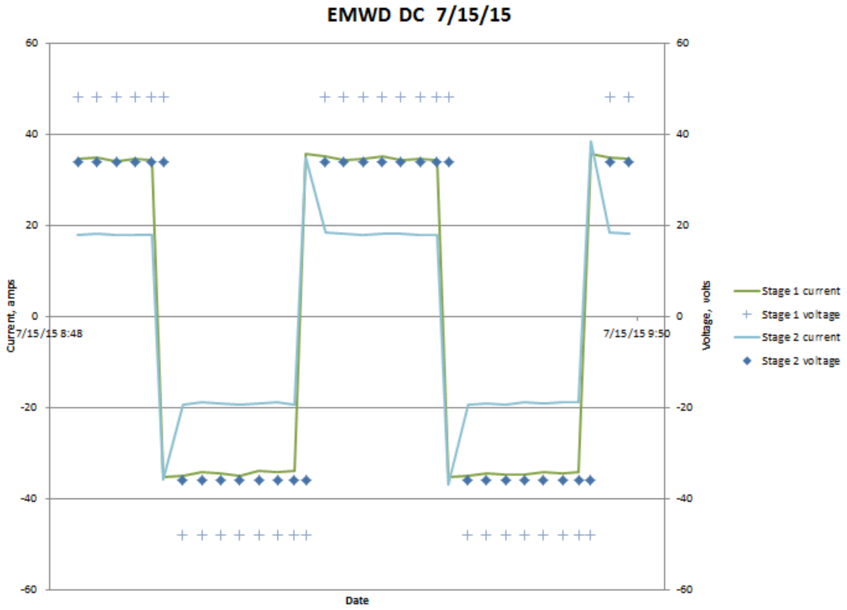

Figure A43. Voltage and amperage 7/15/2015.

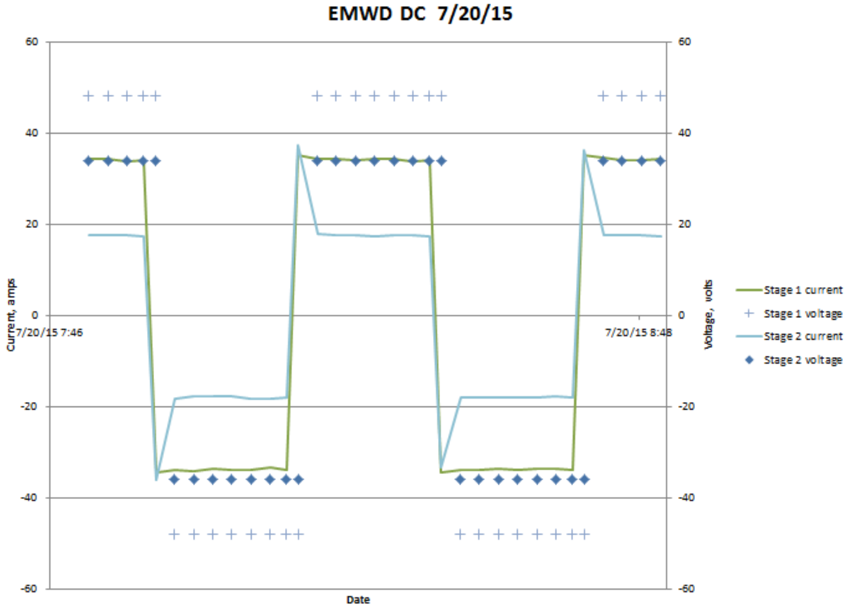

Figure A44. Voltage and amperage 7/20/2015.

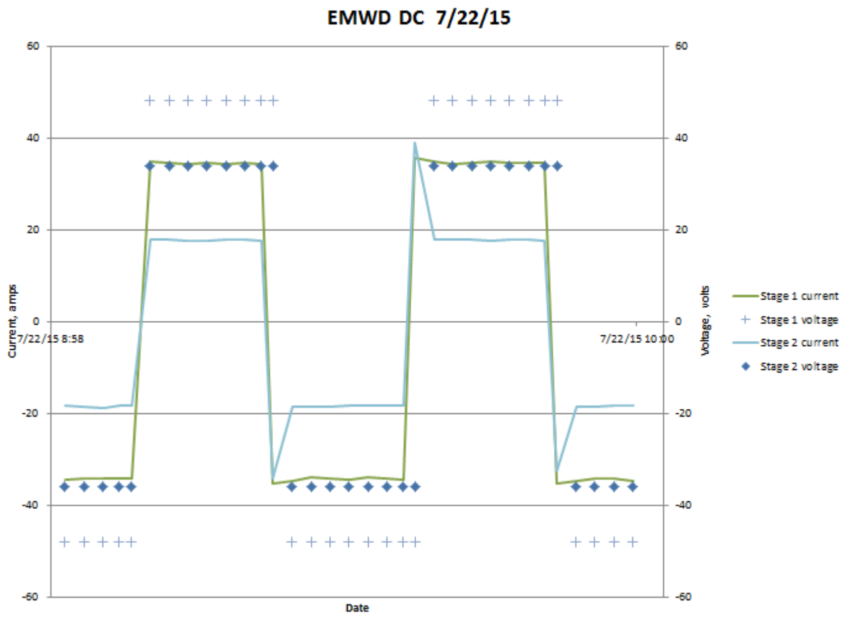

Figure A45. Voltage and amperage 7/22/2015.

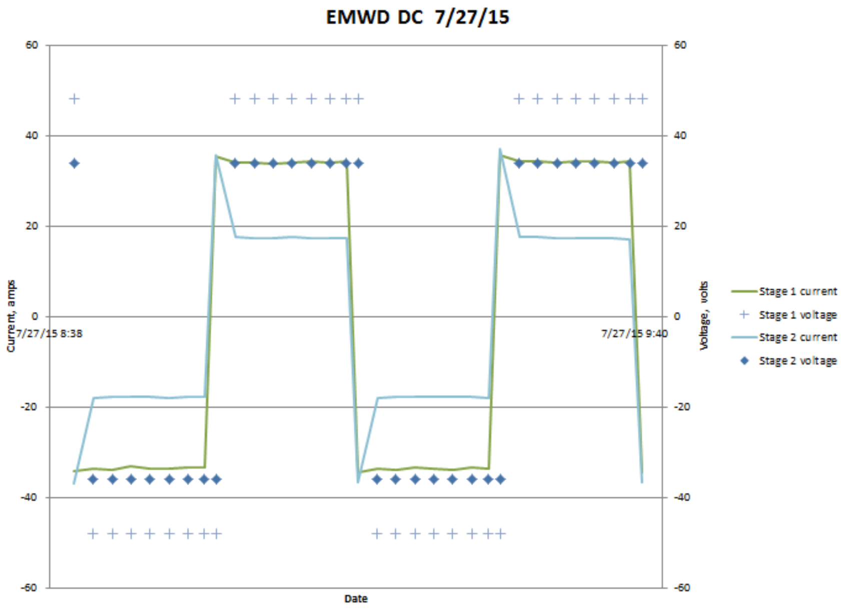

Figure A46. Voltage and amperage 7/27/2015.

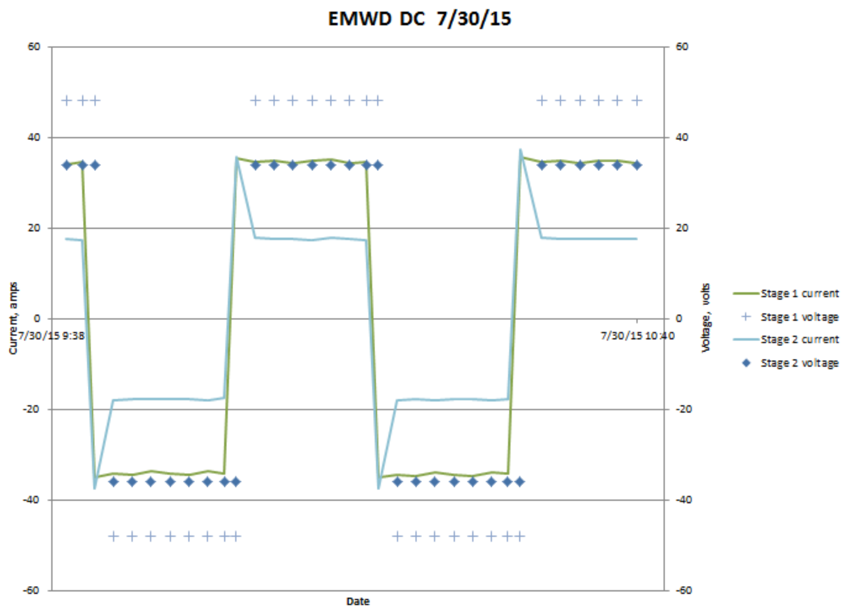

Figure A47. Voltage and amperage 7/30/2015.

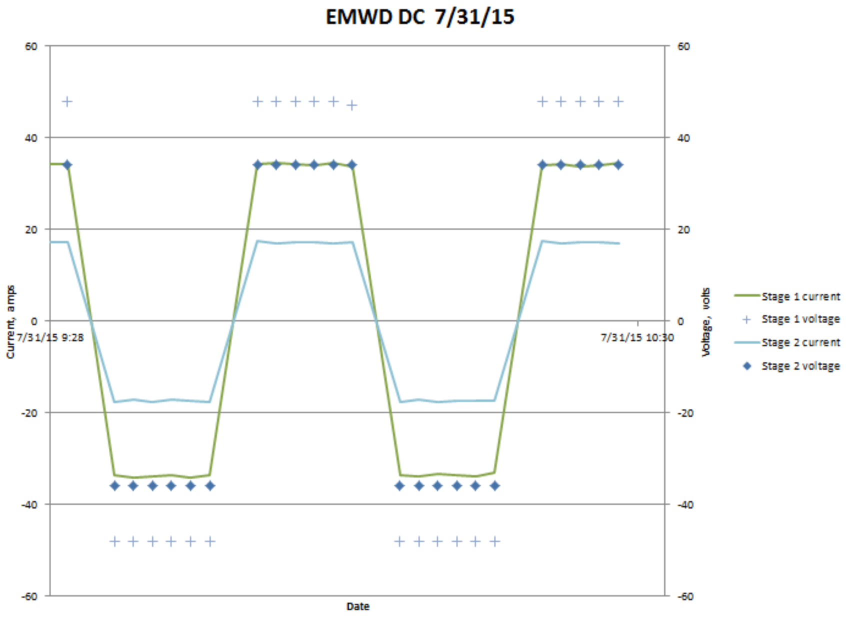

Figure A48. Voltage and amperage 7/31/2015.