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Innovative Electrocoagulation Membrane Pretreatment with Vacuum Assisted Electro-Distillation Concentrate Management for Cooling Tower Blowdown Recovery

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6. AUTHOR(S)
Eric Dole, PE, PSAP – Primary Investigator | Water and Energy Practice Lead | Garver
Robert E. Boysen, PE – Membrane SME | Process Mechanical Design Center Lead | Garver
Guillermo Delgado-Gardea, PhD, PE – QA/QC | Sr. Process Engineer | Garver
Mike Mickley, PhD, PE – QA/QC | Mickley & Associates
Kamyar Sardari, PhD, PE – QA/QC | Process Engineer | Garver
Jessica Johnson – Pilot Plant Operator | Garver
Tyler LeClear – Pilot Plant Operator | Garver

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Garver, LLC
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village, CO 80111

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14. ABSTRACT

The goal of this project is to show the “proof of concept” treatment efficacy and energy intensity (kWh/kgal) of an electrocoagulation (EC)/clarification (CLAR)/microfiltration (MF)/reverse osmosis (RO)/vacuum assisted electro-distillation (VAED) concentrate crystallization system in treating the high total dissolved solids (TDS) commercial cooling tower blowdown. The 600-ton evaporative cooling system is located at Red Rocks Community College’s (RRCC) Main Campus Building located in Lakewood, Colorado. Keeping the 1,200 gpd of 2,200-2,500 mg/L TDS cooling tower blowdown out of the sewer and returning the permeate back to the cooling tower will:

- Help improve water quality of the tower
- Reduce chemical consumption of the cooling tower system because of efficacy in removing calcium magnesium, silica, and bacteria/viruses
- Reduce the make-up water potable water demand to replenish the blowdown
- Eliminate the TDS loading to local sewer shed from blowdown

15. SUBJECT TERMS

Desalination, electrocoagulation, ultrafiltration, granular activated carbon, reverse osmosis, concentrate treatment, vacuum assisted electro-distillation, cooling tower blowdown, ohmic heating, water reuse, zero liquid discharge, onsite sodium hypochlorite generation

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Desalination and Water Purification Research Program
or Pitch to Pilot

Innovative Electrocoagulation Membrane Pretreatment with Vacuum Assisted Electro-Distillation Concentrate Management for Cooling Tower Blowdown Recovery

Prepared by:

Eric Dole, PE, PSAP | Garver, LLC
Robert E. Boysen, PE | Garver, LLC
Kamyar Sardari, PhD, PE | Garver, LLC
Guillermo Delgado-Gardea, PhD, PE | Garver, LLC
Mike Mickley, PhD, PE | Mickley & Associates, LLC

Prepared for:

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Acronyms and Abbreviations

%	percent
C	Centigrade
CIP	Clean-In-Place
CLAR	Clarification
CST	Concentrate Storage Tank
EC	Electrocoagulation
EDR	Electro Dialysis Reversal
FMA	Failure Mode Analysis
FO	Forward Osmosis
GAC	Granular Activated Carbon
HCI	Hydrochloric Acid
HDPE	High Density Polyethylene
HRT	Hydraulic Residence Time
ICP	Inductively Coupled Plasma
MF	Microfiltration
MFR	Moderate Foulant Removal
NAWI	National Alliance for Water Innovation
OFR	Optimal Foulant Removal
ORP	Oxidation Reduction Potential
PFAS	Polyfluoroalkyl Substances
PVC	Polyvinyl Chloride
PWI	Powell Water Systems Inc
Reclamation	Bureau of Reclamation
RO	Reverse Osmosis
RRCC	Red Rocks Community College
SCR	Silicon-Controlled Rectifier
TCLP	Toxicity Characteristic Leaching Potential
TDS	Total Dissolved Solids
TOC	Total Organic Carbon
TSS	Total Suspended Solids
UF	Ultrafiltration
U.S.	United States
VAED	Vacuum Assisted Electro-Distillation
VFD	Variable Frequency Drive
ZLD	Zero Liquid Discharge

Measurements

BTU	British thermal unit
cm	centimeter
°F	degree Fahrenheit
gpd	gallon per day
gpm	gallon per minute
gph	gallons per hour
Hg	inches of mercury
kgal	kilogallon
kWh/kgal	kilowatt-hour per thousand gallons
lbs	pounds
mg/L	milligram per liter
µg/L	microgram per liter
ppm	parts per million
psi	pounds per square inch

Variables

D_p	pore diameter,
D_{int}	interpore distance
D_{den}	pore density
U	anodization potential

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Executive Summary

Cooling towers can be a significant contributor to salt loading in the local watershed. Most industrial and commercial cooling systems utilize evaporative cooling towers, especially in arid regions. The water that doesn't evaporate in the evaporative cooling process cascades down the cooling tower internals into the supply tank, while the higher quality water evaporates. This causes the minerals and salts to "upcycle" in the cooling tower supply water. The upcycling of minerals, especially hardness and silica, causes scaling of the heat exchanger surfaces and a significant decrease in cooling efficiency. For concrete cooling towers, chloride upcycling can cause structural damage to the concrete.

To maintain the proper water chemistry in the supply water, chemicals such as antiscalants and biocides are added. Then, a portion of the water, known as blowdown, is wasted to the local sewer, while fresh water from the local potable water supply is added to the cooling tower system to replenish the blowdown to the sewer and make-up the water lost to evaporation. The high-salinity blowdown can present treatment challenges for water resource recovery facilities and ultimately increase the salt loading to receiving waterbodies.

In 2019, Garver (Denver Office) proposed researching an innovative process train to recover the saline blowdown from commercial and industrial cooling towers under the Bureau of Reclamation's (Reclamation) Pitch to Pilot desalination research funding. Garver was one of seven teams awarded part of the $1.5M Pitch to Pilot grant in November 2019. The grant aimed at researching disruptive desalination technologies that consume less power than what is currently offered on the market. Garver teamed with several entities that played a significant role in project success, such as Red Rocks Community College (RRCC), Dr. Mike Mickley, and our industry partners identified in the logo at the end of this article that collectively donated over $480,000 of in-kind equipment, instrumentation, and labor.

Garver proposed researching a 1.5-gallon per minute (gpm) electrocoagulation (EC), clarification (CLAR), microfiltration (MF), ultrafiltration (UF), granular activated carbon (GAC), reverse osmosis (RO) with vacuum assisted electro-distillation (VAED) concentrate crystallization process train to recover the saline cooling tower blowdown in a zero liquid discharge configuration, preventing the high-salinity wastewater from entering the sewer shed. The water quality of the treated permeate and distillate was analyzed to determine if it can be repurposed for cooling tower make-up dilution water to help balance water chemistry within the recirculation loop. Because of time constraints, the reuse capability for irrigation of the adjacent greenhouse was not evaluated.

The goal of this project is to show the "proof of concept" treatment efficacy and energy intensity kilowatt-hour per thousand gallons (kWh/kgal) of the previously mention process train in treating the high total dissolved solids (TDS) commercial cooling tower blowdown. The 600-ton evaporative cooling system is located at the RRCC Main Campus Building in Lakewood, Colorado. Keeping the 1,200-gallon per day (gpd) of 2,200-2,500 microgram per liter (mg/L) TDS cooling tower blowdown out of the sewer and returning the permeate back to the cooling tower will:

- Improve the water quality of the cooling tower.
- Reduce chemical consumption of the cooling tower system because of efficacy in removing calcium, magnesium, silica, and bacteria/viruses.
- Reduce the make-up water potable water demand to augment the blowdown.
- Eliminate the TDS loading to local sewershed from blowdown.
- Serve as a model for future brackish water membrane desalination systems.

If the pretreatment train was effective at removing the common foulants prior to RO, the system should perform at a higher permeate specific flux without damaging the membranes and allow for high recoveries, while maintaining a salt removal efficiency greater than 99.4 percent (%).

Powell Water Systems Inc. donated their 1.5 gpm EC unit for the duration of the project. EC is an alternative method of adding a metal coagulant without adding a salt counter ion by using direct current and metal electrodes. Under a direct current, the metal anode dissolves into the water, which adds the metal coagulant to the water while also generating H+. OH- is generated at the cathode while also generating hydrogen gas under certain conditions. A ferric hydroxide or aluminum hydroxide “sweep” floc is formed depending on what blades are used. Most of the sweep floc settled in the custom “underflow” clarifier; however, a 20-micron roughing filter followed by a 0.2-micron cartridge filter was used to remove the pin-floc from the EC supernatant that did not settle in the clarifier.

Since a true UF cartridge filter, capable of achieving nominal particle removal less than 0.1-microns was not available, a 0.2-micron MF cartridge filter was used to remove particles less than 20-microns but greater than 0.2-microns, prior to entering the RO feed pump. The GAC was tested to remove total organic carbon (TOC) or any free chlorine prior to RO elements; however, during the beginning of testing it was discovered that the GAC was not needed to remove TOC or free chlorine because the EC was effective at doing this, so the GAC was removed. The final, main treatment train consisted of EC + CLAR + MF + RO. During the FMA, the 20-micron filter was replaced with a 5-micron cartridge filter.

The 2-stage, 2.5-inch diameter DOW FilmTec^{TM} SW30-2540 seawater RO membrane system initially used in the pilot was replaced with a 3-stage system during the first month of operation after a test run significantly exceeded the maximum flux rate, causing the membranes to delaminate and fail. Each array offered 29-ft^2 of surface area and was rated for 99.4% salt rejection at 800 psig at 700 gallons per day (gpd) or 0.486 gpm of permeate flow for a feed concentration of 32,000-mg/L TDS at 77^{o}F, resulting in specific flux of 0.030 gfd/psi. Adding the 3rd stage resulted in increased recovery and allowed us to operates at an average specific flux of 0.046 gfd/psi on the lower salinity blowdown. Throttling the 1st stage permeate valve was required to maintain proper crossflow velocity and higher recoveries in the absence of multiple arrays in the first two stages and with the lower salinity blowdown (as compared to seawater).

However, using three arrays in the first stage followed by two in the second stage and one in the last stage should have increased the recovery and specific flux while decreasing the energy intensity. Furthermore, utilizing brackish water RO membranes versus seawater RO membranes would likely further improve the performance metrics previously stated. Each stage only had one array due to budget restrictions.

The concentrate from RO was sent to a holding tank where the VAED system would treat saline residual flow from the RO in 8 gal batches. The VAED system is a patented distillation process designed to treat the much higher salinity wastewater from the regeneration of salt-based ion exchange resin. The process combines ohmic heating and vacuum distillation into one unit process. It flashes high TDS waters and condenses the steam into low TDS distillate, leaving an 80% (by weight) salt slurry. Modifications were made to the water softener spent brine treatment VAED system, donated by Salt Miner LLC, to try and accommodate the anticipated 3 to 5 gallons per hour (gph) of concentrate from the RO. The standard unit is sized for 1 gph of softener spent brine that can exceed 120,000 mg/L TDS.

Rockwell Automation power quality monitors were installed on each component of the process train to read kilowatt (kW) and kilowatt-hour (kWh). Endress Hauser Pico Mag meters and in-line water quality analyzers were installed throughout the process to trend total flow, instantaneous flow, conductivity, pH, oxidation reduction potential (ORP), and temperature. Together, this instrumentation allowed our team to trend kWh consumed per thousand gallons (kgal) treated and the treatment efficacy in a real-time environment.

Throughout the field testing, parameters were adjusted in the EC process train such as hydraulic residence time (HRT), electrode material, electrode jumping configuration, amperage density, pretreatment chemistry for pH adjustment and downstream filtration micron removal size. The specific flux and the permeability of the RO system was challenged based on the various EC configurations, with some offering lower concentrations of foulants entering the RO than other configurations. Electrode configuration, amperage density, re-seeding of salt slurry, and condensation energy adjustments were made to the VAED system throughout testing.

As shown in table 1, a compilation of all seven tests and 23 grab sample events, the main treatment train consistently met water quality goals and resulted in high TDS removal efficacy without fouling the RO membranes, with total hardness concentrations feeding the RO averaging 429-mg/L as $CaCO_3$. The body of the report will disseminate the results of each configuration with respect to water quality trends recoveries and specific flux. It will also reveal results of the failure mode analysis (FMA) that was initiated after the official field tests were completed. The pilot system was operated three to five hours a day for three days a week between August 2021 to November 2021.

Table 1.—Water quality

	10th Percentile			Average			90th Percentile		
TARGET CONSTITUENT	Raw Water	Filtered EC Supernatant	Permeate	Raw Water	Filtered EC Supernatant	Permeate	Raw Water	Filtered EC Supernatant	Permeate
TH as CaCO3 (mg/L)	700	174	0.100	743	429	1.38	810	723	4.86
pH	8.1	7.9	7.1	8.3	8.8	8.8	8.7	9.3	9.9
TDS (mg/L)	1683	1515	5.00	1819	1686	11.4	1992	1966	21.2
Silica (mg/L)	15.6	0.30	0.3	17.7	1.39	0.3	19.4	3.24	0.3
TSS (mg/L)	5.0	5	5	5.3	12.5	5	5.3	17.8	5
Total Phosphate (mg/L)	0.11	0.05	0.05	0.16	0.06	0.05	0.19	0.05	0.05
TOC (mg/L)	5.79	4.84	0.5	6.38	5.18	0.51	7.34	5.46	0.5
*ORP (mV)	180	-142	-183	194	-112	-125	211	-70.0	-61.2
*Temperature (°C)	15.9	13.2	12.9	18.1	17.0	17.0	19.9	21.2	21.3
Total Coli (mpn/100 mL)	1.0	1.0	1.0	1.18	1.67	1.0	1.0	1.0	1.0
OPERATING PARAMETER	10th Percentile			Average			90th Percentile		
Energy Intensity (kWh/kgal)	39.6			44.9			51.4		
Feed Pressure (psi)	312			346			386		
Permeate Backpressure (psi)	30			37			46.2		
Permeat Flow (gpm)	0.717			0.797			0.864		
Perm Flux (gfd)	11.9			13.2			14.3		
Specific Flux (SF) (gfd/psi)	0.040			0.046			0.050		
MFR Recommended SF (gfd/psi)	0.030			0.030			0.030		
Concentrate Flow (gpm)	0.80			0.88			0.97		
RO TDS Removal Efficiency	99.7%			99.3%			98.9%		
% Recovery 3 stg RO	43%			48%			52%		
**% Recovery Overall	49%			54%			59%		

*As trended through in-line analyzers **@ 400 ppm TDS w/ Blend

Initially, after each RO run during the pilot test, the membranes were forward flushed with 8 to 10 gallons (< 1% of processed flow) of less than 20 mg/L TDS permeate until the concentrate conductivity was within 20% of the permeate feed flush conductivity, thereby indicating the membranes were effectively flushed or recovered. Later in the research, the membranes were operated without a flush between runs and the electrocoagulated MF filtrate was left in the RO system for up to one month during pilot phase testing and up to three months during FMA. The FMA is described in more detail later in the report.

For the VAED system to operate with the existing electrode configuration on the lower TDS concentrate seen during the initial phases of testing (as opposed to higher strength spent brine it was originally designed for), it needed the addition of rock salt to bring the TDS from what was leaving the 3rd stage as concentrate between 2,200 mg/L and 4,000 mg/L, up to approximately 56,000 mg/L TDS. Otherwise, it would take much longer to upcycle the salt to allow the electro-distillation process to occur efficiently. One of the benefits of the VAED system is the higher the conductivity, the quicker the flash occurs because of ohmic heating principles[1]; therefore, counteracting the boiling point elevation phenomenon that occurs with most commercial crystallizers. Three sample sets were submitted for a lab analysis, while the conductivity, pH, and ORP sensors provided additional water quality results.

Notable Trends During Pilot Phase:

1. Increasing the pH with sodium hydroxide from 8.6 to between 9.7 and 10.3 resulted in total hardness removal efficacy prior to the RO between 59 to 80% in the presence of antiscalants.

2. Typically, no more than 170-mg/L to 256-mg/L as CaCO3 total hardness (TH) is recommended in the RO feedwater without proper pre-treatment with antiscalants or pH adjustment to prevent irreversible membrane fouling. An average TH concentration feeding the RO for all test runs was 428 mg/L as CaCO3 with a maximum TH of 735 mg/L as CaCO3 in Test No.3, which was ambient pH, concentrate recycled to EC at a 1-minute hydraulic residence time. No fouling or process degradation was experienced.

3. The electrocoagulated floc, otherwise known as e-floc, was significantly diminished after approximately 360 to 540 gallons of permeate was produced because of scaling of the electrodes and required frequent clean-in-place (CIPs) with 34% hydrochloric acid (HCl) if a higher hardness removal efficiency was desired ahead of the RO.

4. Because of the high hardness concentrations in the blowdown, polarity reversal frequency of the EC unit needed to be increased from once per minute to four times per minute to extend runtime between CIPs and prevent scaling. EC runtimes between CIPs were increased by approximately 30% after increasing polarity reversal frequency.

5. The CIP HCl was recycled between eight to eleven times until the pH increased above three, where it was neutralized and disposed. Per the Powell Water Operation and Maintenance (O&M) Manual for newer EC units (than the 17-year-old EC unit used in this pilot) the spent EC CIP solution can be processed through the EC unit. However, the spent CIP solution may require dilution with potable water to decrease the conductivity to within limits of the EC unit's silicon-controlled rectifier (SCR), or the unit will shut down to prevent damage to the electronics.

6. In all tests, EC depressed the average ORP from 194 millivolt (mV) to -112 mV indicating a reducing condition and the absence of oxidizers that would harm the membrane.

7. Regardless of pH or HRT, the EC unit removed between 83% to 98% of silica, even at pH ranges between 9.7 and 10.3, where amorphous silica has a solubility concentration greater than 800-mg/L at 22° centigrade (C).

8. The GAC system did not provide TOC removal and was not a benefit to the process train. In fact, an increase in total suspended solids (TSS) was noticed in the GAC runs likely because of sloughing, so this configuration of testing was immediately stopped so as not to potentially foul the membrane.

9. The custom fabricated underflow clarifier did allow some floc carry over at a 1-min HRT (1.5 gpm) causing noticeable build-up in the 20-micron cartridge filters but not on the 0.2-micron filters. There was no e-floc carryover with the 3-min HRT (0.5 gpm).

10. A 1-minute EC HRT with ambient pH, traditionally showed the poorest hardness removal efficiency. However, when tested in the FMA, it resulted in one of the best performing 3-stage RO runs with respect to percent recovery and specific flux.

11. The VAED system removed 98 to 99% of the TDS for three lab analysis dates where the raw TDS was 186,060 mg/L, 100,013 mg/L and 103,531 mg/L at an energy intensity between 620 to 1,745 kWh/kgal to create up to 5 gph of high-quality distillate. The lower energy intensity at the higher TDS concentrations confirms the ohmic heating effect on the system efficiency.

12. Higher conductivity brine feed to VAED caused a more aggressive boil in the flash chamber (because of ohmic heating) than anticipated. This resulted in the 5,200 BTU/hr chiller not being able to keep up with condensing the steam at the same rate it was generated, which results in a loss of vacuum and increased boiling point. To prevent melting the polyvinyl chloride (PVC) flash chamber, the power supplied to the heating element across the brine circuit was reduced, which decreased the distillation efficacy. A new, higher capacity chiller was not in the budget.

13. A 12,000 British Thermal Unit (BTU) air conditioner and used radiator was integrated into the cooling loop of the VAED recirc water to help enhance the cooling effect in the heat exchanger in the absence of the chiller. This allowed the unit to attain steady state operation by increasing the condensation rate, so it matched the steam generation rate. Under this configuration the VAED attained a 5-gph distillation rate between 100ºF and 120ºF at a 21-inch Hg to 23-inch Hg vacuum on greater than 180,000 mg/L TDS salt seeded concentrate.

Findings of the Failure Mode MA had profound impacts on the findings of this research. The concept of the FMA was to stress the system after all the field piloting was completed with the goal of "breaking the system" or finding the weak point of the treatment train. This was done by: 1) operating under the worst performing test run configuration under higher pressures and specific flux; 2) continuing to operate infrequently without initiating membrane pickling routines during downtime exceeding 5-days; 3) not performing a CIP on the EC blades despite inability to produce a dense e-floc.

Notable Trends During Failure Mode Analysis (FMA):

1. After field testing was complete in November 2021, the research team began the FMA by recirculating the concentrate to the EC in a proprietary manner with no chemical pretreatment to the RO, only EC + CLAR + MF. The feed pressure was increased to 600-psi +/- 20 psi from an average of 346-psi realized during the pilot test.

2. After producing between 480 to 500-gallons of permeate (over weeks of start stop operation and no pickling of the membranes), the feed pressure rate of gain started to increase from 600-psi and the permeate flow decreased. This initiated a 10 to 15 gallon concentrate purge to the VAED system. In the initial phases, this was followed by a 10 to 15 gallon permeate flush. However, later in the FMA only 10-15 gallons of concentrate purge was initiated and the permeate flush was not incorporated after the concentrate flush and the specific flux and TDS removal efficiency consistently recovered to values detailed in table E-1.

3. Throughout the FMA, the electrocoagulated, 0.2-micron filtrate was left in the RO elements without adding sodium bisulfite preservative for one to three months. Standard operating procedures called for a 15 gallon concentrate purge upon start-up. If the specific flux would not recover to the values in table ES-1 after the concentrate purge, a 10 to 15 gallon permeate flush consistently recovered the membranes to values in table ES-1.

4. There was not a laboratory water quality analysis during the FMA, only surrogate measurements from inline analyzers as previously mentioned.

5. The TH concentration feeding the RO was estimated to be at least 4 times that of the pilot phase described in table ES-1 based on the conductivity increase from the concentrate recycle as read from the inline conductivity meters.

6. An EC CIP was only initiated one time during the FMA prior to a pilot trailer tour to show attendees the abilities of EC in generating a dense, heavy floc.

7. In all tests, EC kept the ORP feeding the RO to below –165-mV indicating a reducing condition and the absence of oxidizers that would harm the membrane.

8. As of January 2024, the system continues to operate in FMA with no pretreatment other than EC at a 1-minute HRT with fouled electrode surfaces. No membrane cleanings occurred, nor was disinfectant or caustic added to pilot system since the pilot began in August 2021. There was no apparent biological growth in the unit processes downstream

of the EC such as the clarifier, clear cartridge filter housings, RO elements, concentrate storage tank and permeate storage tank. Post treatment of the permeate was not performed with respect to pH adjustment or disinfection for all phases of testing,

9. The VAED system continued to remove 98% to 99% of the TDS from the seeded concentrate as indicated by the conductivity tester in the trailer. The rate of distillation consistently reached steady state at 120ºF to 130ºF at a up to 18-gph; however, as the conductivity increased to approximately 65,300-uS/cm, the temperature needed to be decreased because the chiller could not keep up with the flash rate which caused salt laden foam to migrate into the heat exchanger and ultimately into the distillation trough. Salt laden foam in the distillation trough would contaminate the sub-100-uS/cm distillate and require a permeate flush of the distilled water system.

10. The standard 1.5-kW, zinc coated resistive heating element showed no signs of corrosion or scaling after sitting submerged in 100,000 to greater than 200,000 mg/L brine for over 30 months, while being periodically operated 16 to 24 hours a month.

Our hypothesis for this “antiscalant-like” phenomenon of foulant prevention is likely due to the EC. The EC process floods the water with negatively charged electrons which causes a negative ORP and a slight negative charge on the water. Since the polyamide RO membranes are also negatively charged, a repulsion effect of common divalent, positively charged scaling components was noticed. There may also be an induction time impact that the EC introduces that prevents scaling of common foulants. This phenomenon aligns with the findings of Dr. Kamyar Sardari’s research identified in Section 1.7 – Previous Research. Future research should be conducted to prove this concept.

Proof of concept was achieved. The treatment train consistently and effectively was operated in two configurations that resulted in enhanced reverse osmosis operation and high-quality permeate water, which met the water quality goals for use back in the cooling tower recirculation loop with no post treatment.

In the optimal foulant removal (OFR) mode, the EC/CLAR/MF train removed target RO foulants such as total hardness and silica. This mode incorporated a 3-minute HRT, a 9.7 to 10.3 pH and achieved up to 80% total hardness removal (in the presence of residual antiscalants used in the main cooling tower treatment system) and 98% silica removal. The specific flux ranged between 0.04 to 0.05-gfd/psi and achieved 50% to 54% recovery with no fouling, compaction, or delamination. An average pressure of 346-psi was applied in these tests. In the OFR mode, it was assumed the RO concentrate was suitable for sodium hypochlorite generation, after being distilled in the VAED system, because there was insignificant hardness and silica in the salt slurry due to it being removed with EC earlier in the process.

In the moderate foulant removal (MFR) mode, conducted during the FMA, the EC system was operated at a stressed condition by increasing the feed pressure to 600-psi +/- 20 psi, with no pH adjustment and 100% concentrate recirc mode until the membranes experienced performance degradation. Under MFR operation the concentrate purge likely had high hardness concentrations in the concentrate, making it unsuitable to be used as a feed stock for onsite sodium hypochlorite generation after processed with the VAED system. The RO membranes did not appear to foul because they are still operating under this mode of operation. In fact, the highest recovery (97%) was consistently achieved in the MFR mode with a non-optimal RO design, allowing the VAED system to close the ZLD loop. In the MFR mode, the WQ was trended through the previously mentioned in-line monitors in the trailer and not through lab tests.

In summary, the process train could be operated in an OFR mode where the residual salt slurry from the VAED system may be suitable for use as a feedstock for onsite sodium hypochlorite generation. The sodium hypochlorite may be used back in the cooling tower recirc water loop to help promote a circular economy but requires further testing to confirm. Under this mode of operation, the pretreatment costs will likely be higher because of the longer EC HRT and/or caustic addition; however, it could potentially render a beneficial reuse feedstock as a residual. The same operating pressure implemented in the FMA was not attempted in the OFR mode, but it is likely to assume increased recoveries would be realized if it were. In the MFR mode, the concentrate leaving the RO would likely have too high of total hardness concentrations to make it usable as a feedstock for on-site sodium hypochlorite generation, because it was not removed in the EC step. This mode of operation may still be desirable because of the reduced operating costs of pretreatment and the higher recoveries achieved.

With automation, enhanced RO recovery and increased cooling efficiency of the VAED system, the EC/CLAR/MF/RO with VAED treatment train will continue to achieve zero liquid discharge (ZLD) treatment for the blowdown from the 600-ton cooling tower at 1.5-gpm at significantly lower energy intensities. Automation of the treatment train would also allow for longer, unstaffed runtimes so the system can be tested for more than five to six hours at a time. More research into the removal mechanisms, optimization of full-scale design constraints, capital costs, and operational costs would be beneficial to the advancement of this technology as a solution to manage the salt loading from cooling towers, brackish water desalination, and potentially seawater desalination applications. Additionally, in future phases, more research should be focused on the electro-antiscalant phenomenon, and the fate of the residuals generated from the process. More budget for lab analysis would be helpful as well.

1.0 Introduction

1.1 Project Background/The Problem

It's easy to contaminate our precious water supply with salt water, but removing it isn't as easy as shown in figure 1. Salt comes from a variety of sources and understanding the salt loading source in a watershed sets the stage for salinity management. Often there are best management practices that can reduce salt loading to the watershed, thereby reducing treatment intensity.

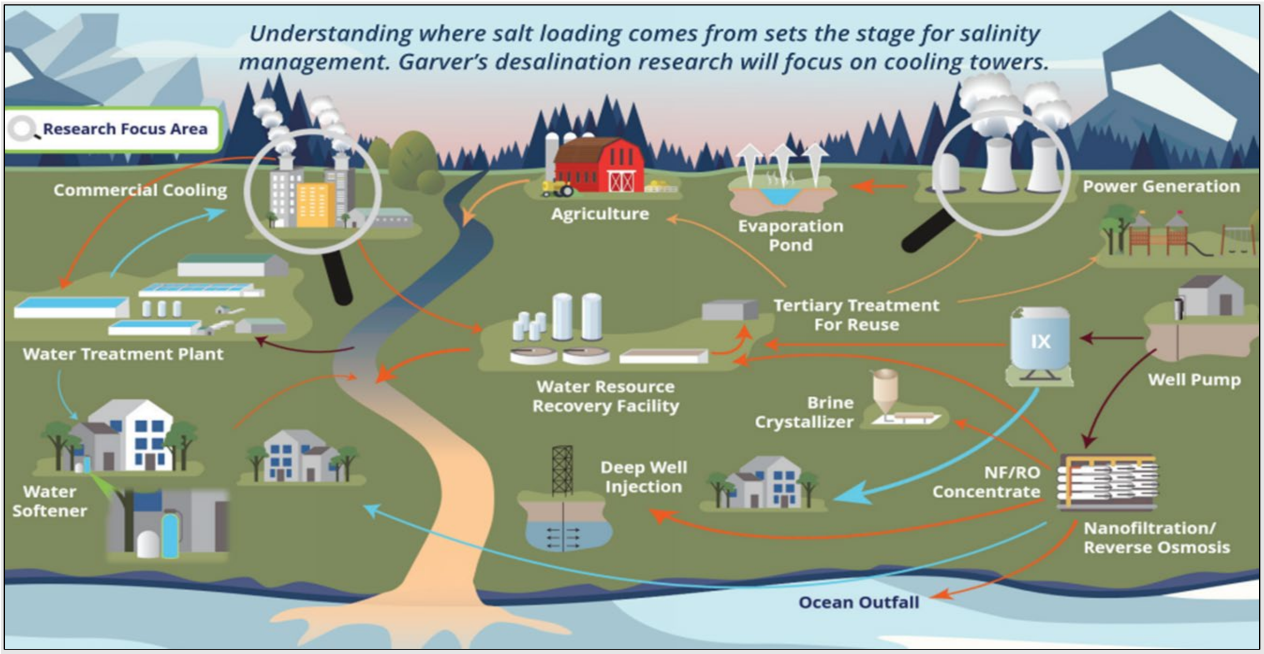


Figure 1.—Watershed salt loading.

Cooling towers can be major contributors to salt loading in the local watershed. Most industrial, commercial, and power generation cooling systems use evaporative cooling towers, especially in arid regions where they are very effective at cooling. The water that doesn't evaporate in the evaporative cooling process cascades down the cooling tower internals into the supply tank, for further cycling while the higher quality water evaporates. This causes the minerals and salts to "upcycle" in the cooling tower supply water. The upcycling of minerals, especially hardness and silica, causes scaling of the heat exchanger surfaces and a significant decrease in cooling efficiency. For concrete cooling towers, chloride upcyling can cause structural damage to the concrete.

To maintain proper water chemistry in the supply water, chemicals must be added, such as antiscalants and biocides, plus a portion of the water must be wasted out of the system as the conductivity increases to a predetermined value. Fresh water from the local potable water supply is added to the cooling tower system to replenish the waste flow leaving the system and to make up for the water lost to evaporation. This saline water wasted to the sewer is known as blowdown. For the majority of industrial and commercial cooling towers, the blowdown is wasted to the local sewer system; however, for cooling towers used in power generation evaporation ponds are typically used to store the blowdown. The higher the cycles of concentration (COC) the cooling tower operates, the higher the blowdown salinity is.

This high-salinity blowdown may present treatment challenges for water resource recovery facilities and ultimately increase the salt loading to receiving waterbodies. According to the Central Arizona Salinity Study, "A study performed by the Phoenix Water Conservation Department concluded that 15 % of all water produced goes to cooling towers. These units use water cycled through the system to cool the air through the process of evaporation. The water is discharged to the sewer system when the salt concentration gets too high and starts impacting the cooling process. Standard practice calls for concentrating brine three to five times before discharging to the sewer". However, in unique situations such as the Palo Verde Generating Station (PVGS) in West Phoenix, the evaporative cooling towers used to cool the 4.2 gigawatt electric (GWe) nuclear reactors operate at 28 to 32 COC's which meant the water entering the facility at 1,000 mg/L total dissolved solids (TDS) is blowndown to the evaporation ponds when it reaches 28,000 to 32,000 mg/L TDS. PVGS is the second largest nuclear reactor in the country and the only one in the world cooled by 80 million gallons per day (MGD) of treated wastewater from the 91st Avenue Wastewater Treatment Facility.

The 600-ton evaporative cooling system that provided the blowdown for the pilot research was located at Red Rocks Community College's (RRCC) Main Campus Building located in Lakewood, Colorado. According to the cooling tower operator, the system is operated at three to five COC, and the blowdown is activated when the conductivity meter reaches the set-point corresponding to 2,500-parts per million (ppm). The system will bleed between 1,200 and 1,500-gallons of water in a batch configuration, while simultaneously introducing lower TDS make-up water from the local potable water supply until the conductivity drops to 2,000-ppm. It's estimated that the blowdown from this one 600-ton cooling tower contributes between 25 to 31 pounds per day (lbs/day) of TDS to the local sewer system, while also consuming 1,200 to 1,500 gallons per day of potable water.

In 2019, Garver (Denver Office) proposed researching an innovative process train to recover the saline blowdown from commercial and industrial cooling towers under the Bureau of Reclamation's (Reclamation) Pitch to Pilot desalination research funding. Garver was one of seven teams awarded part of the $1.5million (M) Pitch to Pilot grant in November 2019 aimed at researching disruptive, desalination technologies that consume less power than what is currently offered on the market.

Garver teamed with several entities that played a significant role in project success such as RRCC, Dr. Mike Mickley, and our industry partners identified in apendix A that collectively donated over $480,000 of in-kind equipment, instrumentation, and labor.

Garver proposed researching a 1.5-gallon per minute (gpm) electrocoagulation (EC)/clarification (CLAR/ microfiltration (MF)/ultrafiltration (UF)/granular activated carbon (GAC)/reverse osmosis (RO) with vacuum assisted electro-distillation (VAED) concentrate crystallization system to recover the saline cooling tower blowdown in a zero liquid discharge configuration, preventing the high-salinity wastewater from entering the sewer shed. Early in the research, the team discovered a cartridge UF was not readily available that met the required design specifications, and the GAC was not needed; therefore, these two-unit processes were removed. Figure 2 shows a process flow schematic of the treatment train and how it was integrated into the existing cooling tower system.

The water quality of the treated permeate and distillate was analyzed to determine if it can be repurposed for cooling tower make-up dilution water to help balance water chemistry. Because of time constraints, the reuse capability for irrigation of the adjacent greenhouse was not evaluated, nor was the quality of the VAED residual waste brine determined if it could be repurposed such as for feedstock in the chlor-alkali market.

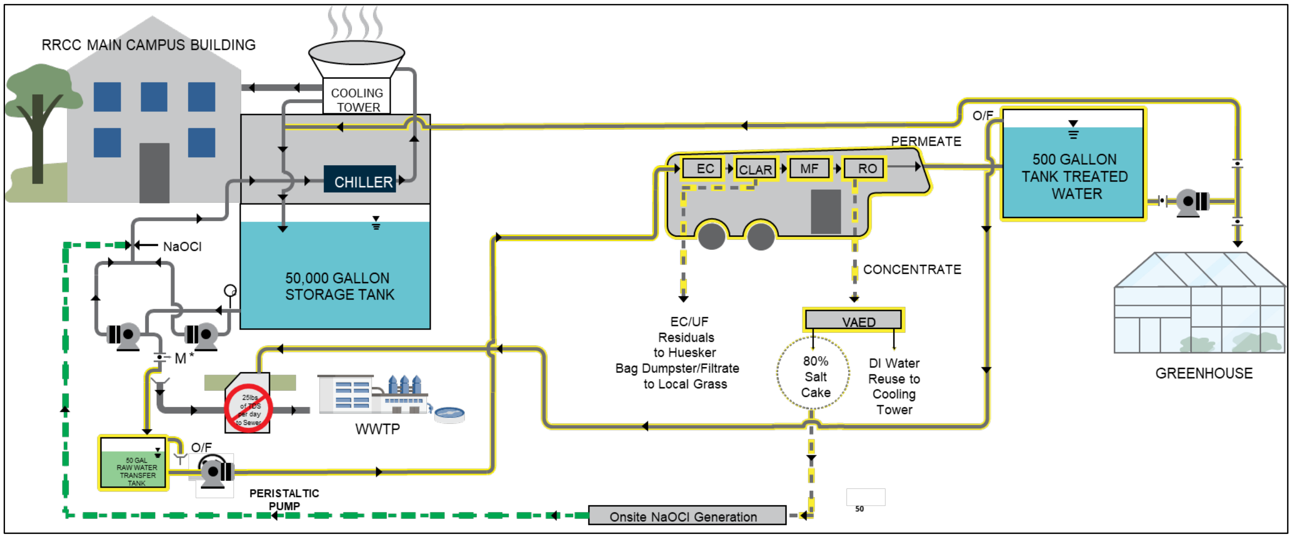


Figure 2.—Garver's overall process flow diagram.

1.2 Purpose

The goal of this project is to show the "proof of concept" treatment efficacy and energy intensity (kWh/kgal) of an EC/CLAR/MF/GAC/RO with vacuum assisted VAED concentrate crystallization system in treating the high total dissolved solids (TDS) commercial cooling tower blowdown. If the EC precipitates the fouling constituents effectively, it's expected that the RO system can be operated at a higher recovery allowing more permeate to be returned to the cooling tower recirculation loop, while less flow of a higher TDS concentrate can be sent to the VAED system (U.S. Pat No. 8,273,156).

The VAED system combines ohmic heating and vacuum distillation into one unit process that achieves the equivalent of three to five multiple effects depending on the configuration and concentration of the brine it is treating. Ohmic heating is when an alternating current is passed through a conductive material such as brine or concentrate, resulting in a rapid rise in temperature of the brine and a faster steam generation rate than traditional radiant heating systems.

It's anticipated the permeate from the main treatment train and the distillate from the VAED system would be of such high quality, with respect to TDS, hardness, silica, phosphorus, and no biological activity, that it could be recycled back into the cooling tower water supply to help increase the COC without any post treatment. The residuals from the main treatment train should be bound as iron oxide solids and pass toxicity characteristic leaching potential (TCLP) tests. The super saline residual brine stream from the VAED system is anticipated to be of high enough quality to be used as a feed stock for the chlor-alkali market, such as sodium hypochlorite generation.

1.3 Objectives

Keeping the 1,200 gpd of 2,200-2,500 milligams per liter (mg/L) TDS cooling tower blowdown out of the sewer and returning the permeate back to the cooling tower will meet the following objectives:

- Help improve the water quality of the tower resulting in increased COC.
- Reduce chemical consumption of the cooling tower system because of efficacy in removing TDS, calcium, magnesium, silica, and bacteria/viruses from the recirculation water stream.
- Reduce the make-up water potable water demand.
- Eliminate the TDS loading to local sewer shed from blowdown.

The following is a list of research related objectives with this project:

- Accurately document all research methodology so it can be reviewed by industry peers.
- Document removal efficacy of target constituents for various treatment configurations.
- Document energy intensity of each major treatment train component (EC, membranes, VAED).
- Hypothesize the removal mechanisms that influence the results.

- Develop Class 5 Level Estimates on a 1-mgd RO system from three major equipment manufacturers utilizing the raw water quality and EC treated water quality (that fed the RO skid) to determine the impact of EC on RO costs.

Lessons learned from this research can be translated to other municipal and industrial brackish water desalination opportunities.

1.4 Project Drivers

It is well-known that membrane desalination is critical for potable water supply in arid coastal regions and land locked areas relying on brackish groundwater; however, management of the concentrate stream is costly and has dire environmental impacts if not addressed properly. Membrane desalination is also essential in many manufacturing industries, power generation and industrial/commercial cooling applications.

The timeliness, uniqueness, and disruptiveness of this treatment train to have a profound effect on other brackish and ocean desalination applications are main project drivers for this research. If successful, this research will provide a high-level treatment train model that can be further researched and refined to improve the operation, energy intensity, and capital costs of a commercialized, zero liquid discharge RO desalter with concentrate treatment.

1.5 General Methodology

The unique aspect of our research conducted in this pilot project focuses on the “electrification” aspects of our primary treatment train with EC and brine residuals treatment train with VAED. The term electrification of water treatment has recently been coined by the National Alliance for Water Innovation (NAWI) when describing some of their A-PRIME C initiatives (Autonomous, Precise, Resilient, Intensified, Modular, Electrified, and Circular).

According to NAWI, the term electrified means when electricity is used in the actual treatment unit process to drive chemical and physical reactions, or in the use of generating chemicals like chlor-alkali market, and not simply providing electricity to power equipment needed to push/pull water or air through a unit process.

Figure 3.—RRCC students conducting EC bench test.

The EC pretreatment to the membrane system was the workhorse of the main process train. Without the EC in place to precipitate the primary membrane foulants (for example: hardness, silica, and total suspended solids) the RO system would have likely irreversibly fouled causing poor recovery and excessive energy intensity. Therefore, our team performed extensive bench testing, prior to pilot testing, with 19 different tests to determine which configuration resulted in the best hardness removal efficiency as a surrogate for the best water quality feeding the RO system in the pilot. Because pH plays a significant role in hardness removal, it was also trended alongside hardness data points. Figure 3 shows the RRCC students conducting the bench tests under the supervision of the EC manufacturer and Garver's principal investigator. To align with budget, an Orbeco-Hellige MC500 Multiparameter Colorimeter was used during the bench tests for hardness tests along with a calibrated HACH Pen Pal pH meter so multiple lab analysis did not have to be submitted to the lab.

The top seven tests of the 19 various configurations tested, that had the highest hardness removal efficiency, were replicated in the pilot train over the course of the 5-month pilot testing. It was decided to use a combination of iron and aluminum blades in the pilot. The results for 23 sampling events were collected from nine different locations across the pilot train and trended in Section 3.

The VAED system could not be bench tested because there was not a bench test unit available. However, various configurations of the VAED system were tested during the pilot phase by adjusting operation variables such as the salt seeding concentration, electrode configuration, and cooling mechanism.

1.6 Challenges/Remedies

Garver was awarded the project in November 2019, and a contract was not executed until September 2021 because of the COVID 19 Pandemic. During this unforeseen delay in the project, several items impacted our budget and schedule, some of which were related to project logistics and others had a process impact on the research. Here are the following challenges along with how our team addressed them:

- **Challenge 1** – The company that was going to build the VAED system, Lyons Filter Inc., could not hold the space in their production line to assemble a new unit capable of treating the 3-4 gph required from the RO concentrate.
    - **Remedy 1** – Salt Miner LLC donated the 1-gph prototype and coordinated shipment from San Carlos, California to RRCC (Lakewood Campus). The cost to ship and retrofit the unit to accommodate a 3-4 gph capacity was less than the $10,000 cost estimate budgeted for Lyons Filter to build a new one.
- **Challenge 2** – A membrane manufacturer was going to donate a 1.5 gpm 2-stage seawater RO skid with all the pumps and ancillary support equipment. They asked for a $21,000 refurbishment fee once the system was decommissioned and returned. After informing them of our award, we were told they already donated the unit they were allocating for this project to another academic research project, and it was no longer available.
    - **Remedy 2** – Our team found a replacement 1.5 gpm 2-stage seawater RO system with the same pumps and ancillary support equipment for $6,200 that was available for immediate shipment. Minor modifications were required to meet the specifications of the pilot.
- **Challenge 3** – Piping, tanks, instrumentation, chemical feed pumps, and raw water transfer pumps were needed to equip the pilot and the $150,000 budget was not enough to cover the inflated costs because of COVID.
    - **Remedy 3** – Several entities donated over $480,000 of in-kind equipment, instrumentation, and labor to equip the pilot trailer to meet the design intent. Please refer to Appendix A for a list of all the companies that contributed to the pilot trailer along with the itemized monetary contribution of each project partner.
- **Challenge 4** – The lab analysis for all the target analytes was $672 per sample times five sample locations per event times 23 events for a total of $74,000. This is roughly half of the project budget.

  - **Remedy 4** – CO Analytical reduced the cost per sample by 10% and Endress + Hauser donated in-line analytical instrumentation (pH, oxidation reduction potential [ORP], temperature, conductivity, and flow) at six locations in the process train that could be used to trend treatment performance. Refer to figure 7 for locations of the sample tap locations and analyzer locations in the process train.

- **Challenge 5** – Lack of pressure transmitters to trend transmembrane pressure, specific flux, and net driving pressure.

  - **Remedy 5** – E+H Pico Mag meters and pressure gauges on feed, concentrate, and permeate allowed specific flux and net driving pressure to be calculated periodically, but on-line trending was not possible. Pressure gauges before and after the MF filter cartridges and UF filter cartridges revealed when to change out the filter because of plugging when pressure differential exceeded 15 psi. Clear polyvinyl chloride (PVC) housings also allowed operators to visually inspect the state of the filter solids loading.

- **Challenge 6** – The original 2-stage RO was delaminated and irreversibly damaged by accidentally exceeding the flux requirement during the first month of operation. Also, the 2-stage configuration could not get the desired recovery while still meeting the permeate flux requirements of 0.030-gfd/ft^2 at the 2,000 to 3,000 ppm TDS versus the 32,000 ppm TDS seawater concentrations it was designed for.

  - **Remedy 6** – Replacement membranes were purchased, and an additional 3rd stage was added to allow higher recovery. Permeate back pressure between 25 and 40-psi was induced on the 1st-stage as required to allow proper crossflow fluid velocity and flux in the absence of multiple arrays in the 1st and 2nd stages. During FMA the permeate backpressure increased to 60 to 80-psi.

- **Challenge 7** – The default polarity reversal frequency in the Powell EC unit is set to reverse polarity in one-minute intervals. Regardless of the test configuration, after approximately 20 to 30 seconds the amperage started to drop significantly because of passivation of the cathode from the elevated total hardness and silica concentrations.

  - **Remedy 7** – The polarity reversal frequency was changed in the control panel to reverse polarity every 15-seconds before the passivation inhibited treatment. This prolonged the Clean-in-place (CIP) cleaning frequency substantially.

- **Challenge 8** – After field testing was concluded in December of 2021, the space heaters in the trailer tripped and allowed freezing conditions in the trailer. This caused damage to the filter cartridges and two of the pH/ORP probes. After careful inspection and testing of the RO system, it was determined that the elements, boost pump, and tubing were not damaged.

  - **Remedy 8** – A new wooden platform was designed to hold a parallel MF/UF train to promote lower head loss and double the filtration area. This was when the FMA started and a 1-minute EC HRT with no pH adjustment and no permeate flushing occurred from the beginning of FMA until now, January 2024.

- **Challenge 9** – Pressure transmitters and data loggers were not available to trend head loss across the prefiltration cartridge filters, nor were they available to trend specific flux for the RO for each of the 3 stages.

  - **Remedy 9** – Properly sized, glycol filled pressure gauges were placed before and after each pretreatment filter array and on the concentrate line. The pressures were manually recorded throughout the daily operation, along with several other pertinent field data, such as flow, pH, ORP, conductivity, and temperature.

1.7 Previous Research

To our knowledge, there are three previous research projects that have similar alignment with the research goals set forth in this project. They are as follows:

- According to Powell Water Systems Inc., (PWI) the application of electrocoagulation (EC) for membrane pretreatment has been successfully employed for one of their confidential clients in Texas at a natural gas power plant. At this facility, the 1,000 gpm EC, UF, RO cooling tower blowdown treatment system operated at 90% recovery and was configured in a zero liquid discharge configuration where the concentrate was sent to onsite evaporation ponds. The permeate was sent back to the boiler feed water.

- Dr. Mike Mickley, our research team member, extensively researched the Powell EC System for the Office of Naval Research in 2004 (Contract No. N00014-04-C-0027) as pretreatment to RO on naval ships where the concentrate was discharged into the ocean and the permeate was used for boiler feed water. Dr. Mickley identified some deficiencies with the EC system. The most prevalent deficiency was the passivation of electrodes because of high concentrations of common foulants like hardness and silica that significantly reduced runtimes between CIPs. PWI has since addressed these deficiencies in their newer units with control logic to allow more frequent polarity reversal as needed and CIP frequency alarms. Despite some of the limitations identified in his research, the conclusion of Dr. Mickley's research states, "Second, the results do clearly indicate the most beneficial application of EC is in terms of providing pretreatment to membrane systems. The use of EC in front of a multi-membrane system of UF/RO or MF/RO has promise to improve performance of the membrane system and to broaden its application to include feedwater having high suspended solids levels".

- Dr. Kamyar Sardari, a process engineer with Garver, was a co-author on a research paper in the Journal of Environmental Chemical Engineering, Volume 6, Issue 4, August 2018, Pages 4937-4944 titled "Electrocoagulation followed by ultrafiltration for treating poultry

> processing wastewater". In this research, Dr. Sardari and his team discovered the following research highlights: 1) Fouling was the main obstacle when operating ultrafiltration with poultry wastewater; 2) A 5-min electrocoagulation HRT significantly reduced fouling on ultrafiltration membrane; 3) Integrated system achieved higher permeate flux and contaminate rejection; 4) Intermittent cleaning led to 50% increase in water recovery in a 7-day period.

This Pitch to Pilot research project is unique in that it will not only incorporate advancements to the PWI EC system that have evolved over the past 18 years, but it will research a technology advancement in concentrate treatment with VAED. VAED is a patented process (U.S. Patent No. 8,273,156) that combines ohmic heating with vacuum distillation to distill water up to 260,000 mg/L TDS to less than 100 mg/L TDS at up to 3-gph capacities.

2.0 Pilot System and Operation

2.1 Existing Cooling Tower Operation

The 600-ton cooling tower chemical treatment program is based on make-up water quality and demand. When evaporation occurs, the cooling tower make-up water meter is activated and sends a pulse over to the chemical feed controller. The controller activates the scale and corrosion inhibitor pump to flow pace the chemical into the recirc water flow. The amount of chemical going into the system is based on a test and adjusted accordingly to achieve the proper amount of inhibitor. A Biocide (liquid bromine) is injected into the recirc water flow based on a weekly program to prevent biological growth. It is activated by a program in the controller to pump the biocide in the system every other day at the hottest time of the day. This is altered depending on winter and summer.

2.2 Pilot Equipment Description

The 1.5-gpm desalination process train researched initially consisted of EC, CLAR, MF, UF, GAC, and RO with VAED concentrate treatment process train. See Figure 4.—3D Rendering of pilot trailer.

Figure 4.—3D Rendering of pilot trailer.
Note: The UF and GAC was omitted from the process train for reasons previously stated.

2.2.1 Detailed Schematic with Sampling/In-line Instrumentation Locations

Piping modifications were made to the existing cooling tower system to allow the saline blowdown to be diverted to a 1,720-gallon swimming pool. See figure 5 for a picture of the blowdown holding tank/swimming pool.

Figure 5.—Picture of blowdown capture pool with members from NAWI/DOE tour on March/23/22.

Figure 6.—Picture of blowdown capture pool during failure mode analysis (FMA) with significant algae growth.

An overflow to the pool was directly connected to the existing floor drain as can be seen in figure 6. In an event the pilot system could not keep up with the blowdown flow rate, it would passively overflow to the sewer. The Flowrox peristaltic pump located 10 feet from the pool in figure 6 was used to flow pace a set flow to the pilot trailer, depending on what EC configuration was being tested, while also serving as a whirlpool-type mixer. Having an accurate flow rate is integral to maintaining accurate treatment variables such as hydraulic retention time in the EC chamber and clarifier. The peristaltic pump allowed us to consistently achieve an accurate flow rate to the treatment process.

As raw water moved through the process train, in-line instrumentation and sample ports were placed at strategic locations as noted by red dots in Figure 7.—Garver's detailed process flow diagram. Rockwell Automation power quality monitors were installed on each component of the process train to read kW and kWh for the EC, membrane, and VAED system. Endress Hauser Pico Mag meters and in-line water quality analyzers were installed throughout the process to trend total flow, instantaneous flow, conductivity, pH, ORP, and temperature. Together, this instrumentation allowed our team to trend kWh consumed per kilogallon (kgal) treated and the treatment efficacy in a real-time environment.

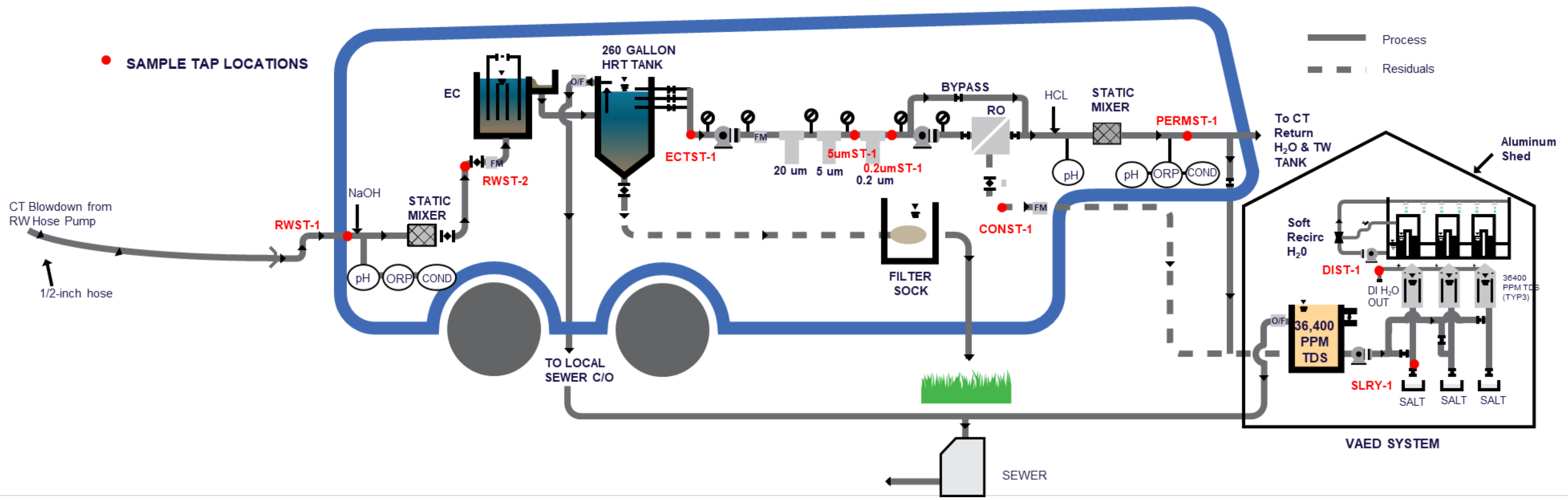


Figure 7.—Garver's detailed process flow diagram.

Throughout the field testing, parameters were adjusted in the EC process train such as hydraulic residence time (HRT), electrode material, electrode jumping configuration, amperage density, pretreatment chemistry for pH and downstream filtration micron removal size, prior to the RO. RO permeate flux and permeability were challenged based on the various EC configurations, and the fact that the 3-stage RO only had one 29-ft^2 element in each stage. Typically, 3-stage RO has multiple elements, or arrays, in the first stages and decreases in number as you increase the number of stages because of the lower flows at higher TDS concentrations. Electrode configuration, amperage density, re-seeding of salt slurry, and condensation energy adjustments were made to the VAED system throughout testing.

2.2.2 Purpose and Description of Each Unit Process

The following section will describe why the unit process was chosen for this zero liquid discharge (ZLD) membrane treatment train and provide a background of the technology, along with pertinent design constraints.

2.2.2.1 Electrocoagulation

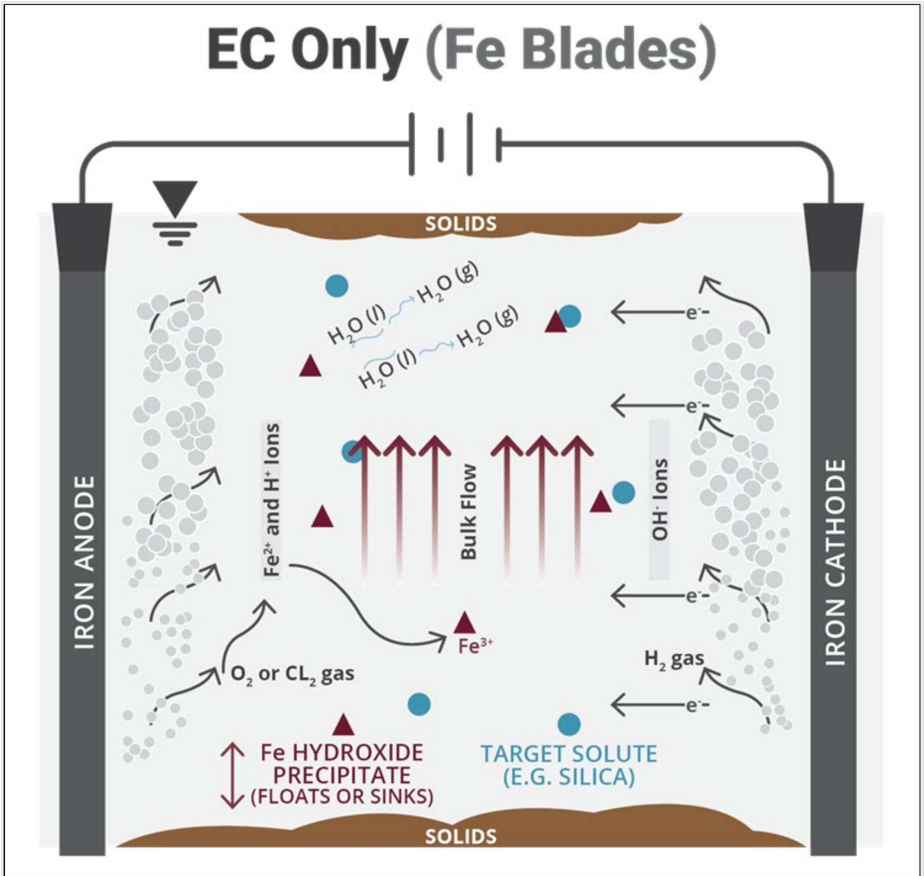
The purpose of EC as pretreatment to a multi-membrane desalination train was to remove the common foulants that would scale and significantly decrease the performance of the RO system. It is well established that EC is an effective treatment process to remove total hardness, silica, total suspended solids, phosphorus and effectively kill viruses and bacteria. See Appendix A-2 for a transmittal letter from Dr. Mya Breitbart with the University of South Florida stating, "In our preliminary experiment, the Powell Electrocoagulation unit reduced all the tested biological agents (including both bacteria and viruses) with greater efficacy than current wastewater treatment practices."

Throughout the pilot testing, the influent pH was either left at ambient between 8.1 to 8.7 or increased with sodium hydroxide to 9.7 to 10.3. Increasing the pH revealed the influence pH adjustment played in hardness removal efficacy in the presence of antiscalants inherently in the blowdown/cooling tower recirc water. Many other parameters were adjusted throughout treatment like HRT, blade configuration and chamber electrode configuration (i.e., chamber jumping) just to name a few.

EC is an alternative method of adding a metal coagulant, like iron or aluminum, without having to add a salt counter ion, such as chloride or sulfate. EC has a long history in water and wastewater treatment dating back to the early 19th Century.[1] This technology was first implemented by wastewater plants in the early 1900's but fell largely out of practice by the late 1930's because of the higher costs of operation related to both the price of electricity and the inefficiency in design. This abandonment proved temporary as renewed interest began an electrocoagulation renaissance in the last decade of the 20th century.[2] Advances in cell construction led to higher removal rates, enhanced treatment capabilities in terms of capacity, and lower costs of both capital investment and operating expenses. These factors have made EC a more viable and economically competitive treatment process when compared to chemical coagulation and flocculation methods.[3]

The EC process consists of an electrolytic cell where DC current is applied across metal electrodes submerged in an electrolyte solution (saline blowdown in this research) which facilitates the flow of electricity. The cell can either be pressurized or atmospheric; however, it is Garver's experience that atmospheric EC cells introduce less operational concerns and perform better with respect to removal efficiencies and energy intensity. The PWI EC unit is atmospheric. Regardless of the EC cell configuration, the exterior housing must be constructed of electrically non-conductive material, such as polyvinylidene fluoride (PVDF), PVC, etc.

Within the EC cell a series of half reactions occur, where oxidation reactions occur at the anode and reduction reactions occur at the cathode. Refer to figure 8 for a functional schematic diagram of the EC process.



$Al\ (s) + 3e- \rightarrow Al^{3+}$ $Fe\ (s) + 2e- \rightarrow Fe^{2+}$ $2H_2O + 4e \rightarrow O_2(g) + 4H^+$	Anodic reactions (Oxidation)
$2H_2O + 2e^- \rightarrow H_2(g)* + 2OH^-$	Cathodic reactions (Reduction)

Figure 8.—EC functional diagram.

EC produces numerous side reactions that enable it to be effective at reducing a wide range of contaminants. One of the primary mechanisms for treatment results from the oxidation at the anode. When applied with a direct current, the anode becomes sacrificial and slowly dissociates metal ions into solution. These metal ions hydrolyze upon injection in the processed water forming metal hydroxide complexes which function as coagulating agents. As with chemical coagulation, the positive metal ions overcome the repulsive forces of colloidal particles allowing the attractive van der Waals forces to dominate, forcing agglomeration.[4] The electrocoagulated floc (e-floc) is instantly formed in the EC reactor at the blade surface and are generally larger, more stable and have higher shear resistance. Passivation or scaling of the cathode blade surface significantly inhibits the efficacy of the EC.

Solids from the EC reactor tend to be hydrophobic in nature, meaning that they inherently repel water, whereas chemical coagulation solids tend to be hydrophilic. The iron hydroxide precipitate (or aluminum hydroxide depending on the material of blades) typically converts to iron oxide over time as it is exposed to air. Refer to figure 9 for an example of the hydrophobicity of solids made from an EC reactor vs non-electrocoagulated solids. In this example, Garver's research staff conducted bench tests on the combined backwash and settled solids from a 16-mgd UF water treatment plant in Lakewood, Colorado utilizing poly aluminum chloride as coagulant. As can be seen, the upper left 11-micron filter paper is the control that had the non-electrocoagulated (suspended solids only) solids accumulated on the paper. The lower right filter paper had electrocoagulated solids accumulated on it, which was about 45% more volume because of the soluble complexes EC removed from the backwash. After two days drying side-by-side in the lab under ambient conditions, approximately 80% less mass of dry solids was left on the electrocoagulated (lower right) filter paper than the non-electrocoagulated (upper left) filter paper.

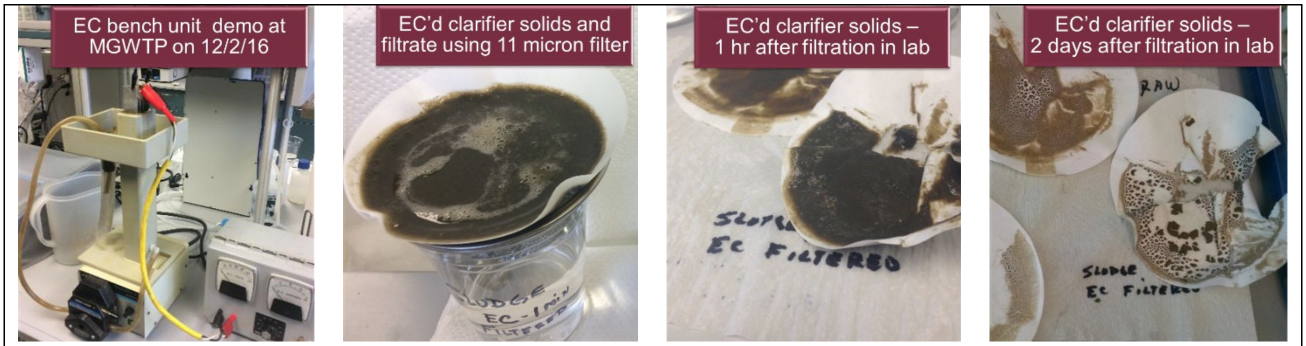
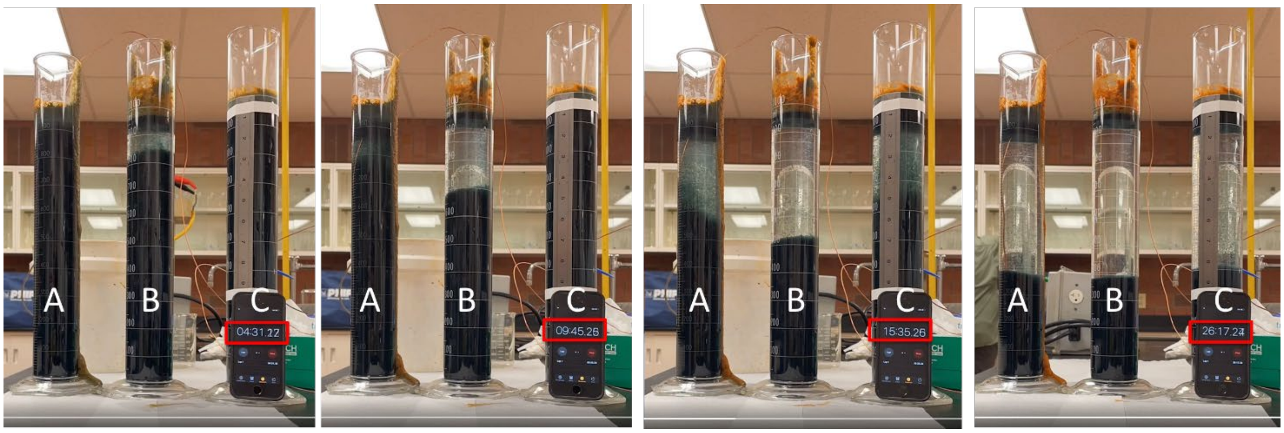


Figure 9.—Hydrophobicity case study of EC solids from UF surface water treatment plant.

Allowing the e-floc to flow into metal pipes, pumps, or tanks will ground the electrons and result in a less stable e-floc. This can be easily demonstrated by allowing an e-floc to settle in a glass beaker. Once settled, the e-floc can be stirred and agitated. In a short amount of time, the e-floc will settle in the same time frame as the initial test. This can be done over and over for several months, and the same settling results can be achieved. Once a copper wire is submerged in the e-floc and inserted into the grounding port in a wall receptacle, the electrons are grounded, and the e-floc will not re-agglomerate the same as before. It takes much longer and resembles the impact of shearing a chemically coagulated floc. Refer to figure 10 of a lab experiment that validates the grounding effect on an e-floc.

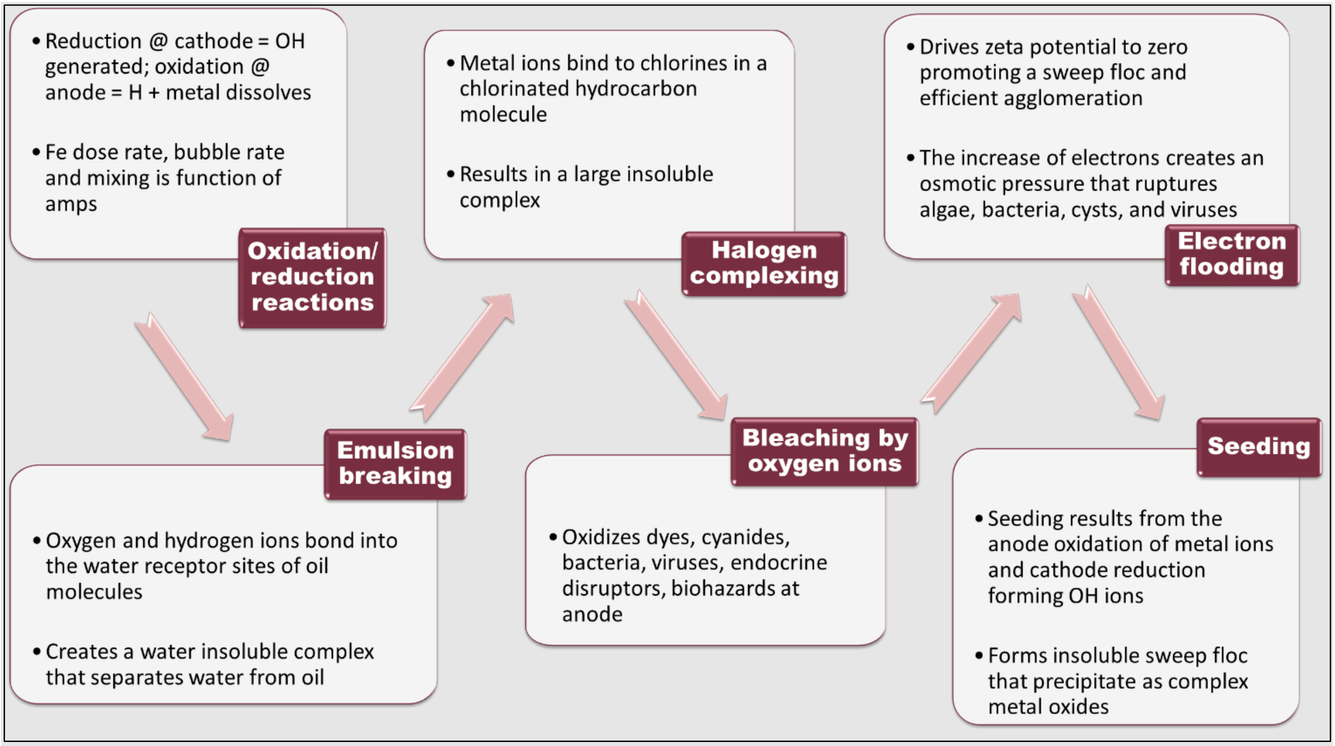


Electro-coagulated municipal wastewater
A - grounded with copper wire to ground on nearby outlet
B - nothing done to e-floc (ie: control)
C - de-aerated with vortex machine

Figure 10.—Grounding Effect on E-floc | Courtesy of Powell Water Inc.

A broad spectrum of pollutants become attracted to this coagulant. The e-flocs created either float (because of the production of hydrogen and oxygen caused by the electrolysis of water) or sink to the bottom of the cell. The EC process is depicted in figure 8. The PWI design used in this research introduces additional air into the reaction chamber to promote floatation of all pollutants so that they may be easily carried into downstream sludge removal processes. The air introduction is especially important to allow e-floc to move on to downstream clarifier, and not settle in the EC chamber, when longer HRTs are required that correspond to lower upflow velocities.

There are six main removal mechanism that occur simultaneously in a properly designed EC cell as described in figure 11.



Oxidation/ reduction reactions

- Reduction @ cathode = OH generated; oxidation @ anode = H + metal dissolves
- Fe dose rate, bubble rate and mixing is function of amps

Halogen complexing

- Metal ions bind to chlorines in a chlorinated hydrocarbon molecule
- Results in a large insoluble complex

Electron flooding

- Drives zeta potential to zero promoting a sweep floc and efficient agglomeration
- The increase of electrons creates an osmotic pressure that ruptures algae, bacteria, cysts, and viruses

Emulsion breaking

- Oxygen and hydrogen ions bond into the water receptor sites of oil molecules
- Creates a water insoluble complex that separates water from oil

Bleaching by oxygen ions

- Oxidizes dyes, cyanides, bacteria, viruses, endocrine disruptors, biohazards at anode

Seeding

- Seeding results from the anode oxidation of metal ions and cathode reduction forming OH ions
- Forms insoluble sweep floc that precipitate as complex metal oxides

Figure 11.—EC removal mechanisms.

These reactions, as well as the others that occur in the process, allow for removal rates of a larger variety of water contaminants than traditional chemical coagulation. EC has been shown to provide modest to highly effective treatment for: arsenic, boron, selenium, suspended solids, phosphate, biological oxygen demand (BOD), hardness, turbidity, Total Kjeldahl Nitrogen (as well as ammonia individually), fluoride, silica, bacteria (E. Coli and total Coliforms), metals (Cu, Ni, Zn, Mn, Pb, Cd, Fe) and other pollutants such as per- and polyfluoroalkyl substances (PFAS).

The electrode materials used in the PWI EC cells are typically either iron or aluminum. These materials possess high valences (aluminum being trivalent, iron being di or trivalent) which contribute to effective destabilization of colloidal charges.[4] They also have the benefit of being inexpensive, commercially available, and relatively non-toxic.[2] While many electrode geometries have been explored, most modern reactor designs utilize flat plates placed in parallel. Along with proven efficiency, it enables cost effective replacement through straightforward fabrication.

The rate at which electrodes are consumed is dictated by Faraday's Law and can be calculated from benchtop testing. This equation is given by:

$$m = \frac{Mit}{nF}$$

Where:

m = mass of metal (Me) electrode that has been dissolved into solution, (g)
M = molecular weight of metal, (g/mol)
i = current, (amps)
t = time duration of reaction, (s)
n = number of electrons or valence state of electrode (No.)
F = Faraday's constant, (C/mol), 96,485 C/mol

The number of electrons participating in the reaction is dependent on the valence of the plate materials. For aluminum, this value would be 3. For iron, the number depends on the pH of the solution as well as the voltage. Higher voltages and pH values tend to favor the production of trivalent ions as opposed to divalent.[4] Increasing the amperage over a set duration result in an increased rate of dissolution of the metal coagulant from the electrode, and hence a higher coagulant feed rate into the solution. The equivalent concept in chemical coagulation is increasing the speed of a chemical feed pump.

Some benefits of EC include a lack of moving parts, ease of operation, faster reaction kinetics, and settling of a more stable sludge than chemical coagulation, no additives to the process that would increase TDS, and the reduction of a significant number of contaminants. The sludge produced is typically lower in volume, more easily dewatered, and capable of passing California Toxic Characteristic Leaching Procedure (TCLP) and Waste Extraction Test (WET). Process integration may lead to reductions of chemical additions as well. Disadvantages include electrode replacement, electricity costs, and reduced efficiency from electrode passivation.[3,4]

Figure 12.—Pictures of 1.5 gpm EC unit in pilot trailer.

The effectiveness of EC in treating challenging contaminants is directly related to several process variables such as, but not limited to, the following:

- **Volts/gap** – The amount of voltage required to make the metal leave the metal blade is about 1.5 volts direct current (DC). A voltage above 1.5 volts simply creates heat in the liquid, consumes more energy, and dissolves the metal blades faster.
    - **Relevance to this research** – The Powell EC System operates at 3 volts/gap.
- **Blade material** – Iron blades should meet American Society for Testing Materials (ASTM) 1011 specification for iron purity and aluminum blades should meet the ASTM 1050 specification for aluminum purity. The blades should be purchased as a commodity and non-proprietary, machined to fit the EC cell configuration. Both iron and aluminum electrodes can be used in the same chamber to address a wider variety of contaminants than either one alone, whereas, in common water treatment coagulation and filtration scenarios, ferric chloride and alum would never be added at the same location in the process train. Aluminum blades remove fluoride well whereas iron blades do not. Iron blades remove iron, manganese, arsenic, chromium, and uranium whereas aluminum blades are not as effective at removing these constituents.
    - **Relevance to this research** – Iron and aluminum blades were used in the pilot testing.
- **Power connection to electrodes** – Electricity takes the path of least resistance. Connecting the DC power source to each blade in an EC cell allows the electricity to flow through the copper wire and not through the less conductive water where the electro-chemical reaction needs to occur at the blade surface. The only time the power connection should be placed closer together is with less conductive water.
    - **Relevance to this research** – During the field testing the conductivity of the blowdown appeared to be ample to allow one chamber, resulting in ~ 5 amperes (amps). After field testing concluded the chamber was jumped and split into two equal chambers during the FMA. This resulted in a larger quantity of floc generated, but also a higher energy intensity in kWh/kgal.
- **Polarity reversal frequency** – Water that has elevated concentrations of hardness or silica will passivate the cathode, or coat it with an electrically non-conducting oxide, at a quicker rate than water with lower concentrations. Once the cathode passivates with scale it reduces the electrical conductivity of the circuit and inhibits the release of metal coagulant from the anode under application of a direct current. Reversing the polarity in the chamber causes the constituent that plates on the blade surface to release off the blade as a solid. More frequent polarity reversal reduces the rate of cathode passivation when higher hardness and silica concentrations are present.

  - **Relevance to this research** – The total hardness in the cooling tower blowdown ranged between 650 – 880 mg/L as calcium carbonate ($CaCO_3$). This high of hardness concentrations caused a rapid decrease in amperage drop because of passivation of the cathode with the default 1-minute polarity reversal duration. The polarity reversal frequency was changed from every one minute to every 15 seconds, which resulted in a marginal amperage drop and longer runtimes between CIPs.

- **Clean-in-place (CIP) protocol** – When the amperage drop rate increases to levels out of specification, an acid CIP is required. Hydrochloric acid is the most common CIP agent; however, some applications require sulfuric acid to dissolve the remaining scale from the cathode surface that the polarity reversal could not accomplish. The Powell Water System recycles the acid until it takes too long to dissolve the blade (pH greater than 3), at which point it is electrocoagulated through the system.

  - **Relevance to this research** – CIP frequency was reduced in the high hardness blowdown water because the polarity reversal frequency was increased from once per minute to once every 15 seconds; therefore, reducing the severity of cathode passivation and downtime due required for a CIP routine.

- **Amp density (amps/in^2 of wet blade surface area)** – The amount of amperage draw is dependent upon the conductivity of the water. If the water is not conductive, then no amperage will be used, and EC cannot take place. EC systems should be designed with adequate wiring and electrical capacity to deliver adequate amperage if needed by a particular water stream.

  - **Relevance to this research** – Powell Water Systems are designed for up to 0.375 amps per square inch depending on the conductivity of the solution. The silicon-controlled rectifier (SCR) in the EC unit is sized to handle this maximum amperage before tripping. This is also why there is a high conductivity chamber and a “normal” conductivity chamber to accommodate the limits of one standard SCR in various conductivity water. The higher conductivity water has a greater distance between the electrode connections than the “normal” chamber configuration, while the wet blade surface area remains the same between the two configurations.

- **HRT (min)** – The hydraulic residence time (HRT) in an EC cell is one of the most impactful variables that directly correlate to the efficacy of EC treatment. More challenging contaminates like selenium and per- and polyfluoroalkyl substances (PFAS) require longer HRTs and/or a chemical catalyst. Most commercial EC units on the market standardize the capacity rating of an EC unit in gallons per minute assuming a 1-minute HRT. For example, phosphorus is typically removed in 15 seconds; therefore, a 600-gpm system can treat 2,400 gpm. Conversely, selenium sometimes requires more than 2-minute HRTs so the same 600 gpm unit can only treat 300 gpm.

  - **Relevance to this research** – A 1-minute and 3-minute HRT was evaluated during the pilot tests. The 1-minute HRT allowed the system to be operated in a continuous flow configuration. During the 3-minute HRT runs, the system required batch operation because the EC flow rate was 0.5 gpm, and the membrane flow rate was 1.5 to 1.6 gpm because of the constant flow of the RO feed plunger pump.

- **pH & Conductivity** – Just as with most treatment systems, pH plays a role in the EC removal efficacy. Often the pH of EC processed water remains near neutral for most waters mainly because of the half reactions that occur simultaneously in the cell where OH^- is generated at the cathode and H^+ is generated at the anode. If the water is highly conductive and dominated by sodium (i.e., spent anion resin brine), the electrocoagulated water will be reduced and tend to have a higher pH because of generation of NaOH- at the cathode. If the highly conductive water is chloride dominated (i.e., spent cation resin brine), the water tends to be oxidized and has a lower pH because of the generation of HCl at the anode. Conductivity of the raw water plays a significant role in the configuration of the EC reaction chamber and dictates: 1) if a standard chamber with two electrode connection points is used; 2) if a standard chamber with electrode jumping is used to split the chamber for non-conductive water; or 3) if a high conductivity chamber is used that has the equivalent wet blade surface area just shorter in height so more blades are installed side-by-side resulting in a longer distance between the two electrical connections.

  - **Relevance to this research** – A standard 1.5 gpm PWI EC unit was used in this research. During the pilot tests the blowdown conductivity was high enough to use a one chamber, non-jumped, configuration where at least 5-amps was used in the chamber; however, if the chamber was split into two equal chambers a much thicker e-floc formed at the cost of higher energy intensity and higher blade consumption. During the FMA, a jumped configuration was used at the initial runs to build a thick e-floc; however, the electrode connections were adjusted to one chamber as the conductivity increased.

- **Post EC settle time** – According to Powell Water, a 2-hour residence time in a clarifier is ideal to allow the e-floc to agglomerate and settle. Additional mixing energy is not required as the e-floc is immediately formed at the surface of the blade once energy is applied. Sometimes the e-floc will rise instead of settle. In these scenarios, dissolved air floatation may be an effective solids separation technique as opposed to gravity settling. Regardless of the settling technique, it is imperative not to have the electrocoagulated floc be introduced into metallic, or electrically conductive, pipe, pumps, mixers, tanks, or instrumentation prior to solids separation, or else the electrons will be grounded. Grounding the e-floc results in an unstable solids, poor filtration, and removal efficacy of targeted contaminants.

  - **Relevance to this research** – See Figure 10.—Grounding effect on e-floc

- **Downstream filtration** – The more effective, or tighter, the filtration, the better the removal efficacy of precipitated solids. Typically, the longer the EC HRT, the larger the e-floc particle sizes will be, assuming the cathode does not passivate or scale in between polarity reversal. The larger the particle size the larger micron removal can be used for filtration. Looser filtration (i.e., larger nominal micron removal size) results in less head loss and less horsepower pumps to push/pull the water through the filter media per unit flow whether filtration is in the form of traditional sand, multi-media, disc, or membranes. On the contrary, shorter EC HRTs tend to result in smaller diameter e-floc particles and require tighter filtration to achieve equivalent target removal efficacies at equivalent hydraulic loading rates. Tighter filtration results in more head loss and higher horsepower pumps to remove the smaller size e-floc.

  - **Relevance to this research** – See Section 2.2.2.3 – Microfiltration, 2.2.2.4 – Ultrafiltration and Section 2.2.2.5 – Reverse Osmosis.

- **Chemical catalyst** – More challenging contaminates such as PFAS, cyanotoxins, selenium and carbon chains shorter than six either require longer process times and/or a chemical catalyst. Adding hydrogen peroxide with iron blades with a one-to-two-minute HRT results in an electro-Fenton reaction that produces a hydroxyl radical. The hydroxyl radical is the second strongest oxidant next to fluorine and will rapidly destroy and precipitate challenging contaminates like the previously mentioned while also effectively killing any viruses, pathogens, and bacteria.

  - **Relevance to this research** – Since one of the main goals of this research was to challenge the effectiveness of EC in removing excessive hardness concentrations upstream of RO, pH adjustment was evaluated prior to EC. Hardness removal is significantly increased as the pH increases to 9.5, which was anticipated, but not to the extent as seen with EC as a catalyst in higher ionic strength water in the presence of antiscalants. For example, hardness removal with 1-minute HRT, pH between 6.5 to 8 and iron blades typically result 40% to 70% removal efficiency. Increasing the pH to 9.5 while holding all other constraints the same, improves the hardness removal efficiency to greater than 90%. It was difficult to attribute how much hardness removal was due to caustic addition versus EC. pH titration tests were performed with 34% sodium hydroxide (NaOH) to determine the proper dose rate in the raw water at a 1.5 gpm and 0.5 gpm flow rate, prior to the EC unit to increase the pH to between 9.7 and 10.

    An E-Fenton or pathogen kill evaluation was not conducted in this research due to budget constraints; however, it would be beneficial to evaluate in future research. The fact that the clarifier, MF or RO elements never fouled with biological activity during the 3-year research project (that is still being operated occasionally in FMA mode) is a qualitative testament to biological kill.

2.2.2.2 Clarification

The purpose of clarification was to allow additional time for the e-floc to form and settle under gravity prior to downstream filtration. Various configurations of clarifiers exist from circular, rectangular, lamella, and solids contact clarifiers. Effective clarification is critical in any treatment process to promote less solids loading on filters, longer filter run times, less backwashing or filter replacement and lower energy intensity (kWh/kgal) for the filtration process.

Our research team designed and built a custom clarifier that was fabricated from a 270-gallon cube tank with a 21-inch diameter center shroud energy dissipator that extended to the bottom of the tank, unlike traditional clarifiers that stop several feet from the surface. At the bottom of the 22-inch diameter center shroud, 12-inch by 8-inch coupons were cut out of the bottom of the pipe in each quadrant to allow the e-floc flow to slowly exit the bottom of the clarifier shroud and get filtered by the settled solids accumulating around the coupons as the cleaner water upflowed towards the floating weir. A floating weir was designed to pull the less turbid supernatant from the top of the clarifier, while solids settled at the bottom. Please refer to Figure 13.—Pictures of underflow clarifier.

Figure 13.—Pictures of underflow clarifier.

The effectiveness of a clarifier in promoting successful solids settling is directly related to several process variables such as, but not limited to, the following:

- **Surface Loading Rate (gpd/ft^2)** – Standard clarifier rates range between 300 to 1,200 gpd/ft2 to promote ample settling velocity gradient.
    - **Relevance to this Research** – The center shroud surface loading rate at the 3-minute EC HRT configuration at 0.5 gpm was 272 gfd/ft^2. As the EC HRT was reduced to 1-minute at 1.5 gpm, the surface loading rate was increased to 815 gfd/ft^2. Additional surface area in the annular space between the circular shroud and the walls of the cube tank increases the overall net surface area and promotes efficient settling.
- **Residence Time (hours)** – Standard clarifier residence times range between 1 to 3 hours. Powell Water recommends an ideal residence time of 2 hours to allow the floc to build and settle.
    - **Relevance to this Research** – Under a 1-minute EC reaction chamber HRT configuration (1.5 gpm), the clarifier experienced a 3-hour residence time and with a 3-minute EC reaction chamber HRT (0.5 gpm) the clarifier experienced a 9-hour HRT.
- **Solids Capture and Removal** – Most commercial clarifiers are sloped toward the center of a circular clarifier and have a rake that turns slowly to direct the solids to a center sump where it is piped to a pump that sends the solids to a dewatering operation.
    - **Relevance to this Research** – The cube clarifier did not have a mechanical rake to remove the solids, as it relied on the 30% sloped bottom to a side drain. Solids would occasionally be drained by opening the 2-inch drain valve and allowing the solids to flow into the cooling tower basement where a 100-micron filter sock was placed on the end of the pipe. Approximately 25 feet of driving head pushed the water through the filter bag where the solids would accumulate.

2.2.2.3 Microfiltration

The purpose of microfiltration was to provide a roughing, outside-in, cartridge filter for the electrocoagulated floc that did not settle in the clarifier prior to the ultrafiltration and reverse osmosis. Microfiltration removes particles in the range of approximately 0.1 – 10 micron. In general, suspended particles and large colloids are rejected well, while dissolved solids pass through the MF membrane. Applications of MF include removal of bacteria, flocculated materials, or TSS. Transmembrane pressures are typically between 3 and 30 psi; however, the cartridge filters used in the pilot had a recommended max pressure differential of 15 psi to limited structural damage to filters.

Figure 14.—Initial 2-stage membrane configuration with 20-micron and 0.2-micron prefiltration.

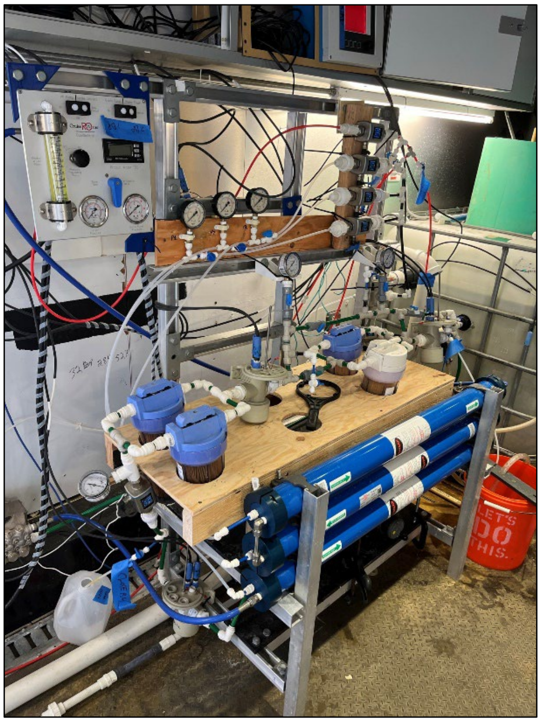


Figure 15.—Modified 3-stage membrane configuration w/ parallel 20-micron and 0.2-micron pretreatment filters.

The effectiveness of MF in filtering suspended solids is directly related to several process variables such as, but not limited to, the following:

- **Flux Rate (gfd)** – Flux rates for membranes are what hydraulic loading rates are for standard filters. It's a normalized metric to track the flow rate over a filtration surface area. Standard microfiltration flux rates in gallons per foot squared per day range between 50 to 100 gfd to promote efficient solids removal and not hydraulically overload the filters.
  - **Relevance to this Research** – Two 20-micron filters removed particles 20-micron and larger that did not settle in the clarifier. The hydraulic loading rate for the 20-micron filters at 1.6 gpm was 356 gfd for the parallel configuration and 712 gfd for a single filter configuration. The hydraulic loading rate for the 0.2-micron filters at 1.6 gpm was 190 gfd for the parallel configuration and 381 gfd for a single filter configuration. Additional surface area in the parallel filters increases the overall net surface area and promotes lower differential pressures through the MF stage to provide ample net positive suction head on the RO feed pump. Despite significantly exceeding the recommended flux rate, the cartridge filter performance was not compromised. It is important to note that the parallel MF filter cartridges were integrated after the field testing concluded in November 2021.

During the FMA the lead filter was replaced with a 5-micron filter instead of a 20-micron filter. All of the pressure loss was across the 5-micron filter and the 0.2-micron polishing filter never had to be changed. The solids accumulated on the 5-micron filter were removed, rinsed with hose water then soaked in permeate for 10 minutes and put back in service over 7 times without experiencing any structural damage that would compromise operation.

- **Boost Pressure (psi)** – Microfiltration requires a driving pressure to overcome head losses through the filter.
  - **Relevance to this Research** – Two self-priming variable speed quad diaphragm pumps were connected for single or series operation. It was anticipated that the tighter 0.2-micron filtration would result in excessive head loss than the 20-micron cartridge filters that came with the skid; therefore, requiring two boost pumps in series. This was not the case, and the second pump was not used. The pumps provided a maximum 45 psi to the microfiltration cartridges and helped provide sufficient suction pressure to the reverse osmosis plunger pump suction to prevent cavitation.
- **Filter Configuration/Material** – MF systems can be configured as crossflow filtration or dead-end filtration. Crossflow filtration can provide efficient filtration provided the flux is not exceeded and ample crossflow velocity is maintained. Crossflow filtration has a constant residual stream and is typically made of polyvinylidene fluoride (PVDF) construction. Dead-end MF filters require a backwash or replacement when the pressure differential is exceeded. Dead-end filters recover all the water during service flow and there is not a constant residual stream; however, there is a waste stream when backwashed. Dead-end filters are typically made of a cellulose media that is pleated to allow greater filtration surface area.
  - **Relevance to this Research** – The 20-micron and 0.2-micron dead-end filter cartridge used in this research were Pentek S1 pleated cellulose sediment filters 10 inches tall and 2.5 inches in diameter. The 20-micron filter offered 3.2 ft^2 of filter area and the 0.2-micron filter offered 6 ft^2. The 5-micron roughing filter used during the FMA had 3.2 ft^2 of surface area and replaced the 20-micron filter used during the pilot phase.
- **Differential Pressure** – As solids accumulate on the surface of the filter, or if the filter surface fouls because of the precipitation of calcium and magnesium complexes or silica, the differential pressure increases and restricts the flow. If the feed pump has a variable frequency drive, the speed of the drive can be increased to a certain degree to overcome the head loss and maintain the proper flux or hydraulic loading rate. Alternatively, a dead-end filter can be backwashed or replaced once the manufacturer recommended differential pressure is reached to prevent structural collapse of the filter and excessive flow restriction.

  - **Relevance to this Research** – The 20-micron, 5-micron, and 0.2-micron dead-end filter cartridge had a manufacturer recommended 15-psi differential pressure max rating. During FMA the 20-micron filter cartridge was removed.

2.2.2.4 Reverse Osmosis

The purpose of RO was to provide semi-impermeable separation of molecules so small effectively removed dissolved salts from the process stream and allow the low TDS permeate to be returned to the cooling tower recirculation loop, while the high TDS concentrate was sent to the VAED system for final treatment. A one horsepower (HP) plunger-style, 1.6 gpm boost pump was used to re-pressurize the 0.2-micron MF filtrate to the 2-stage RO system. Eventually the 2-stage RO was replaced with a 3-stage RO system to increase recovery.

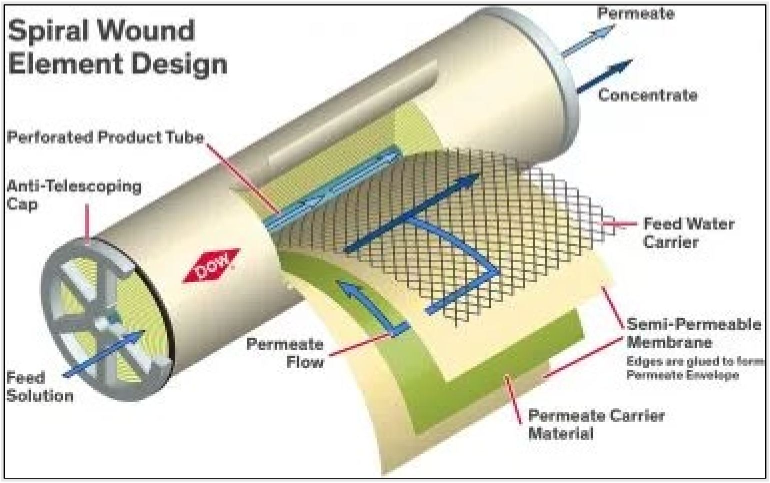


Figure 16.—DOW FilmTec$^{\text{TM}}$ RO module.

RO utilizes crossflow filtration process, whereas the MF system used in the pilot trailer utilized dead-end filtration. The RO process diagram is shown in Figure 16.—DOW FilmTecTM RO module. With a high-pressure pump, feedwater is continuously pumped at elevated pressure to the membrane system. Within the membrane system, the feedwater will be split into a low-saline and/or purified product, called permeate, and a high saline or concentrated brine, called concentrate or reject. A flow regulating valve, called a concentrate valve, controls the percentage of feedwater that is going to the concentrate stream versus through the membrane towards the perforated product tube and exit as low TDS permeate.

The effectiveness of RO in removing salts is directly related to several process variables such as, but not limited to, the following:

- **Recovery** – The percentage of membrane system feedwater that emerges from the system as product water or “permeate”. Membrane system design is based on expected feedwater quality and recovery is defined through initial adjustment of valves on the concentrate stream. The higher the degree of throttling of the concentrate valves, the more backpressure is added on the membranes, and the greater degree of recovery as permeate. Recovery is often fixed at the highest level that maximizes permeate flow while preventing precipitation of super saturated salts within the membrane system.

  - **Relevance to this research** – The percent recovery of the system tested ranged between 55% and 73% depending on the net driving pressure and the efficacy of EC pretreatment. During the FMA the recovery of the EC, CLAR, MF, RO train was increased to 97% consistently under a proprietary configuration, without the use of any chemical pretreatment other than EC.

- **Rejection** – the percentage of solute concentration removed from system feedwater by the membrane. In reverse osmosis, a high rejection of total dissolved solids (TDS) is important.

  - **Relevance to this research** – The percent TDS rejection was approximately 99% under all 7 EC pretreatment configurations, including during the FMA as indicated by in-line conductivity meters on the RO feed, each permeate stage, total permeate and concentrate.

- **Flow** – Feed flow is the rate of feedwater introduced to the membrane element or membrane system, usually measured in gpm or cubic meters per hour (m^3/h). Permeate flow is the portion of the feed flow that penetrates the semi-impermeable membrane as lower conductivity water due to the backpressure placed on the feed flow from the concentrate valve. Concentrate flow is the higher salinity flow that does not penetrate the semi-impermeable membrane and exits the membrane element as the residual waste stream. This concentrate contains most of the dissolved constituents originally carried into the element from the feed source that is not removed in the pretreatment process. It is usually measured in gpm.

  - **Relevance to this research** – The feed flow of the RO system was a constant 1.6-gpm to 1.7-gpm from 1-HP plunger pump. The percent concentrate flow was dictated by the amount of backpressure created by the concentrate valve. The concentrate flow ranged from 0.42-gpm to 0.88-gpm. During the FMA, 100% of the concentrate was recovered in a proprietary manner and occasionally purged from the system; thereby increasing the recovery to consistently 97% without mechanical damage to the membrane. The feed pressure during the FMA was increased to 600-psi +/- 20-psi from the average of 346-psi experienced during the pilot.

- **Flux** – The rate of permeate transported per unit of membrane area, usually measured in gallons per square foot per day (gfd). Each DOW FilmTec^{TM} SW30-2540 seawater RO membrane is 40-inch length x 2.5-inch diameter offered 29 ft^2 of filtration surface area. There was one element in each of the three stages.

  - **Relevance to this research** – Depending on the recovery, the permeate flux ranged between 10 and 22-gfd. Because of the lack of parallel arrays in the first stage, backpressure was applied to prevent maximum flux exceedance and ensure proper cross flow velocity.

- **Specific Flux** – the rate of permeate transported per unit of membrane area per pressure, usually measured in gallons per square foot per day divided by the pressure (gfd/psi). Specific flux is also referred to as permeability.

  - **Relevance to this research** – Specific flux values ranged from 0.04 to 0.05-gfd/psi during the field testing, depending on what configuration the system was operated.

Permeate flux and salt rejection are the key performance parameters of a reverse osmosis process. The flux and rejection of a membrane system are mainly influenced by variable parameters including:

- **Pressure** – With increasing effective feed pressure, the permeate TDS will decrease while the permeate flux will increase.

- **Temperature** – If the temperature increases and all other parameters are kept constant, the permeate flux and the salt passage will increase.

- **Recovery** – Recovery is the ratio of permeate flow to feed flow. In the case of increasing recovery, the permeate flux will decrease and stop if the salt concentration reaches a value where the osmotic pressure of the concentrate is as high as the applied feed pressure.

- **Feedwater Salt Concentration** – As the pressure is held constant and the feedwater salt concentration increases, the permeate flux will decrease and the salt passage will increase.

2.2.2.5 Vacuum Assisted Electro-Distillation (VAED)

The purpose of the VAED system is to flash the high TDS concentrate from RO under lower temperatures and energy intensity than commercial crystallizers. The low TDS distillate can be repurposed in the system, and the salt slurry can be repurposed as a feedstock for onsite sodium hypochlorite generation if it does not have any hardness or silica in it. The onsite sodium hypochlorite could potentially be used to disinfect the recirculation water to help promote a circular economy. A more detailed evaluation of the sodium hypochlorite generation potential will be evaluated in future stages of research.

The RO concentrate in this research project was sent to a holding tank where the VAED system would treat the saline concentrate in 8 gal batches. The VAED system is a patented distillation process for spent brine treatment from water softeners that combines ohmic heating and vacuum distillation into one unit process. The properties of the ohmic heating process –which include the rapid and uniform heating, high energy efficiency, technical simplicity, and low capital and maintenance cost– suggest that it could be a highly competitive heating method for thermal desalination process.[1] Ohmic heating counteracts traditional boiling point elevation concerns typical with commercial crystallizers. With the VAED, the higher the TDS concentration of the brine, the quicker it heats. Coupling this phenomenon with traditional vacuum distillation allows the VAED system to flash high TDS waters at temperatures between 95ºF and 130ºF under a 21-inch mercury (Hg) to 29-inch Hg vacuum (depending on elevation of the unit and local weather) and amp densities ranging between 0.03 amps/in^2 to 0.11 amps per square inch (amps/in^2). The steam rises and condenses in the chilled headspace of the flash chamber turning into low TDS distillate that is collected and stored in the distillate collection trough. An 80% (by weight) salt slurry is left in the bottom of the flash chamber as the super saturated brine level falls to below the bottom of the electrode as steam is flashed. Refer to Figure 17.—Schematic Representation and Picture of VAED System.

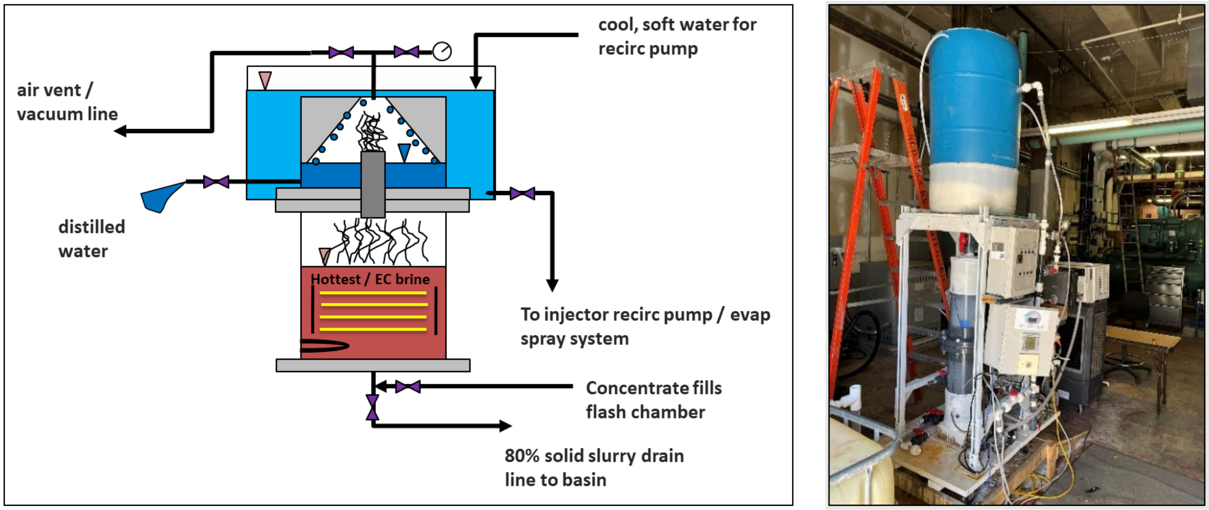


Figure 17.—Schematic Representation and Picture of VAED System.

If the EC pretreatment effectively removed the hardness and silica prior to the RO, then the concentrate make-up would be dominated by high concentrations of NaCl; however, if the EC configuration did not remove the previously mention constituents, then they will also be found in the concentrate feeding the VAED system. The latter would likely render the salt slurry residual from the VAED system not suitable as a feedstock for onsite sodium hypochlorite generation because of the high hardness and silica concentrations. If the hardness and silica were removed from the concentrate by effectively removing it with the EC system prior to the RO, then the spent salt slurry from the VAED system can be used to for onsite sodium hypochlorite generation that can be re-introduced into the cooling tower water recirc loop to promote a circular economy. Modifications were made to the water softener spent brine treatment VAED system, donated by Salt Miner LLC, to try and accommodate the 3 to 5 gph of concentrate from the RO. The standard unit is sized for 1 gph of softener spent brine that can exceed 120,000 milligram per liter (mg/L) TDS.

The effectiveness of VAED in distilling the concentrate into reusable distillate is directly related to several process variables such as, but not limited to, the following:

- **Amp Density (amps/in^2)**- The higher the amp density, the greater the temperature rise rate (Delta °F/Delta seconds), and steam production rate (pounds/seconds), will be. The larger the surface area of the electrode, the more uniform the heating/bubble generation rate will be.
    - **Relevance to this Research** – The VAED system amp density ranged between 0.02 – 0.045 amps/in^2 depending on the conductivity of the brine in the flash chamber.

  Brine Conductivity – According to Ohm's Law (P (kW) = I^2 (amps) R (ohms)), power is dissipated as heat and the quantity of power transferred across the brine circuit is directly correlated to the square of the brine conductivity. As electrons flow through wires, they collide with the ions in the wire which causes the ions to vibrate more. This increased vibration of the ions increases the temperature of the wire. The same phenomenon occurs when electrons flow through water and the water is heated. Given a constant load across the brine circuit in kW, doubling the conductivity results In quadruple the power/heat transferred across the circuit, whereas doubling the resistance only results in doubling the power/heat transferred. The advantage of ohmic heating in thermal desalination is its ability to generate heat. The rate of heat generation will increase as the electrical conductivity increases due to the increasing concentration as the seawater evaporate.[1]

    - **Relevance to this Research**–- The RO concentrate was sent to a holding tank where the VAED system would treat the 2,200 to 4,100 mg/L TDS in 7-gal batches. A salt seed was required to increase the conductivity of the brine to approximately 56,000 mg/L TDS to fit within the design parameters of the demo unit. The salt slurry remaining after the distillate was flashed off and condensed was wasted from the unit three times during the field testing for water quality analysis. Each of the three sampling event resulted in 0.5-L of heavy brine leaving the system mass balance. Other than these heavy brine wasting events, the salt seed was left in the flash chamber to increase the conductivity of incoming 7-gal of concentrate.

      During the FMA, the higher conductivity concentrate was simply transferred to the flash chamber in 5-gallon batches. The conductivity meter in the pilot trailer was used to determine the conductivity of the batches sent to the VAED, which was between 17,000 micro-Sieman per centimeter (uS/cm) to 23,000 uS/cm, or approximately 10,880 mg/L to 14,720 mg/L TDS, respectively. The remaining salt slurry was left in the chamber during the FMA as a "salt seed" for the next batch of concentrate to process.

- **Condensation Efficiency** – Regardless of how efficient the brine is heated, and steam is created, it is important to condense the steam at the same rate it is generated. Otherwise, the pressure in the flash chamber increases causing the vacuum to decrease. As the vacuum decreases the boiling point increases to a point where it can melt the PVC flash chamber and the energy intensity (kWh/kgal) of the unit increases.

    - **Relevance to this Research** – An additional 1,200 BTU per hour (BTU/hr) air conditioner and radiator system was added to the 5,200 BTU/hr chiller to match the rate of steam generation as a result of the ohmic heating in the VAED flash chamber, so steady state operation could be achieved.

      During the FMA, the chiller failed, and the cooling relied only on the recirc water, air conditioner, radiator, and fan. A new heat exchanger helped improve the heat transfer efficiency so the condensation rate could keep up with the higher steam generation rate experienced with the higher conductivity brine in the flash chamber. However, as conductivity increased to approximately 60,000 uS/cm excessive boiling occurred and the condenser could not keep up with the rate of steam generation, so the bubbles would migrate into the pure water side of the heat exchanger. The bubbles contained salt and would cross contaminate the pure water side of the system if the power was not turned down to reduce the height of the bubbles.

- **Electrode Material and Configuration–** The electrode material and configuration can significantly impact the performance of the VAED system. A conductive metal should be used at a thickness that will not allow bending under a vacuum. The electricity will take the path of least resistance so to ensure equal heating across the face of the electrode it is critical to maintain equal spacing between the electrodes. The electrode will dissolve over time and portions blade may become weak causing bending inwards towards the opposite electrode, resulting in a decreased distance between the electrode surface and increased heating at that location only.

    - **Relevance to this Research–** The electrodes in the VAED were aluminum sheets that were 8 in. wide x 1/8 in. thick x 34 in. long. The distance between electrodes was maintained consistent at 5 in.

2.3 Pilot Operation

This section describes the general operation of the pilot system. It is not intended to give a detailed sequence of operation of the pilot trailer. As with most pilot testing, the operation of the pilot trailer was modified throughout the course of the field testing based on lessons learned and optimization opportunities.

2.3.1 Continuous/Intermittent Operation

Because of budget constraints and scope of this research, the pilot trailer was not fully automated to allow continuous operation; however, the individual components of the treatment train were automated, such as the EC and RO system. Communication between the individual unit processes were not automated and relied on the attention of the operator. The duration, and frequency, of a test run was limited to operator availability. As a result, the pilot system was operated three to five hours a day for three days a week between August 2021 to November 2021. The failure mode analysis (FMA) began after the completion of the field testing in November 2021. All water quality data observed during the FMA was from the in-line E+H instrumentation probes, specifically flow, conductivity, pH, and ORP. More details on the FMA are provided in Section 2.3.4.

RO systems are intended for continuous operation and not periodic, start/stop operation as was experienced in this project. This results in a stressed system and results taken from the research can be considered conservative since a fully automated, continuous flow pilot should have better removal efficacy and energy intensity numbers.

2.3.2 Operation of EC – RO Train

Piping modifications were made to the existing cooling tower blowdown loop to allow the blowdown to passively flow to the 1,700-gal swimming pool in the basement of the cooling tower that served as the blowdown collection tank. The blowdown is based on conductivity in the system. It is set to blowdown at 2,000 to 2,500 mg/L and has an in-line conductivity meter that communicates the real-time conductivity reading to the local controller. The blowdown is activated when the conductivity reaches the set-point (2,500 mg/L) and will bleed water down to the lower end of the spectrum (2,000 mg/L). An overflow to the blowdown collection tank allowed blowdown flowrates exceeding our treatment capacity of 1.5-gpm to passively flow to the same floor drain it was previously connected to. A Flowrox peristaltic pump and Rockwell Automation variable frequency drive was connected to the pool to allow precise flow paced delivery of the blowdown to the treatment train located approximately 30-feet higher than the pool. A recirculation line was teed off the discharge line to allow continuous whirlpool mixing in the blowdown collection tank while in operation. This line also served as a pressure relief line in the event a discharge valve in the trailer was closed while the peristaltic pump was still in operation.

Once the blowdown entered the pilot trailer it went through a series of unit processes that can be seen in Figure 7.—Garver's detailed process flow diagram. When the EC unit was ready for operation, it was energized manually at the same time as the peristaltic feed pump was turned on in the trailer. When the EC unit was placed in the auto mode, the inlet valve opened and the blades in the EC chamber were energized. A raw water throttle valve located just before the EC skid was used to fine tune the flow rate to the EC unit as a function of the HRT being tested. For a 1-min HRT the flow rate was set at 1.5-gpm +/- 0.1-gpm. For a 3-min HRT the flow rate was set to 0.5-gpm +/- 0.05-gpm. The 3-min HRT configuration resulted in a batch operation with the RO system since the downstream MF and RO system was rated for a constant flow of 1.6-gpm.

Blowdown was introduced in the bottom of the EC rection chamber and overflowed the chamber weir to the custom fabricated underflow clarifier located 3-feet away. A floating weir allowed the clearer supernatant from approximately 4-inches below the water surface to be pulled through the variable speed, positive displacement, diaphragm MF feed pump and to the cartridge filters. The MF feed pump delivered the clearer supernatant at a 1.6-gpm +/- 0.05-gpm through the 20-micron then 0.2-micron pleated cartridge filter. A 5-micron roughing filter was used during the FMA and replaced the 20-micron filter used during the pilot phase. The 0.2-micron filter was used as the polishing filter for the pilot and FMA phases.

The RO feed pump was a constant speed, positive displacement, plunger-style pump that re-pressurized the 0.2-micron filtrate that ranged between 25-psi to 40-psi, depending on the headloss through the cartridge filters. The RO feed pump delivered a constant flow to the RO system at 1.6-gpm +/- 0.05-gpm regardless of the backpressure placed on the membranes with the manual needle valve on the concentrate line. The membranes were rated for 800-psi for seawater and had a pressure relief valve that was set to 1,000-psi. Since the conductivity of the blowdown was closer to brackish water conductivity concentrations instead of the seawater conductivity the membranes were rated for, the RO feed pressure was set to 340-psi to 400-psi during the pilot tests. The permeate was sent to the 300-gallon permeate collection tank and concentrate was sent to the 300-gallon concentrate collection tank inside the mechanical room located 30-feet away from the trailer. However, during FMA the recovery was consistently increased from 54% to 97% (without fouling, compaction of delamination) for up to 6 hours of operation until the pressure increased to 600-psi to 630-psi. At approximately 480-gallons to 510-gallons of permeate produced, the pressure would increase to 640-psi, at which point 10-gallons to 12-gallons of concentrated concentrate was purged to the VAED system.

During the 3-minute HRT configuration, the reduced flow entering the EC unit required a batch operation where the clarifier would fill to a high-water level while the RO system was off. Once full, the operator manually tuned on the membrane system and pump 1.6-gpm while approximately 0.5-gpm system would continue to flow through the EC unit. This allowed roughly 2.5 hours of runtime to allow the system to stabilize and collect all six sample sets from the various location in the process train. If the test called for pH adjustment, the ambient pH was increased from approximately 8.6 to between 9.7 and 10.3 by manually dosing with 0.25% NaOH at roughly 170-mL/min with the use of the peristaltic chemical feed pump.

Since EC was the main process component being studied to evaluate its efficacy as pretreatment to RO, a significant effort was taken to bench test EC to determine the best performing configuration, while the pilot trailer was being designed and constructed. Bench testing various EC treatment configurations allowed our research team to dial in on the best performing pilot testing configuration with respect to hardness removal. The bench testing impact on pilot testing configuration will be discussed in more detail in Section 3.

There were seven shortlisted scenarios developed from the bench tests where a combination of iron and aluminum blades were used. Refer to Section 4 for a detailed discussion regarding the configuration and results of the various tests performed in the pilot study.

2.3.3 Operation of VAED Train

During the pilot trailer operation, the concentrate from the RO was either sent to the concentrate storage tank (CST) on a continual basis or recycled within the system in a proprietary manner during the FMA until small batches needed to be purged to the CST due to recovery limitations of the system. The purpose of recirculating the concentrate for re-treatment was to increase the raw water conductivity to the system, increase recovery of RO; thereby resulting in a lower flow rate and higher conductivity feed to the VAED system. Because VAED incorporates ohmic heating, the higher the conductivity the quicker the brine was heated. An overflow for the CST was connected to the drain of the adjacent permeate storage tank so anytime the VAED could not keep up with the concentrate flow rate, the permeate was drained and mixed with the concentrate overflow. This allowed the same mass loading of TDS to the sewer as would occur if the pilot was not there.

For the VAED system to operate with the existing electrode configuration on the lower TDS concentrate (as opposed to higher strength spent brine it was originally designed for), it initially needed the addition of rock salt to increase the TDS from the 3rd stage concentrate. The rock salt addition spiked the TDS concentration ranging from 2,200 mg/L to 4,100 mg/L, up to approximately 56,000 mg/L TDS. If the concentrate was not spiked with rock salt, the VAED system would take much longer to upcycle the salt concentration to a conductivity that allowed the electro-distillation process to occur efficiently.

During the pilot test, salt spiked concentrate from the CST was mixed in an initial 8-gallon batch then pumped into the bottom of the flash chamber. Once the concentrate level reached the 8-gallon mark in the flash chamber, the fill process was stopped. Five gallons of permeate, with less than 20-uS/cm conductivity, was added to the water bath at the top of the VAED system and served as the supply water for the venturi injector vacuum system and the chilled water. The venturi feed pump delivered permeate from the elevated bath through the venturi injector and back into the elevated bath in a closed loop. A tee and throttle valve on the discharge of the pump was used to control the flow through the condenser. The condenser was energized at the same time the venturi feed pump was energized. After 25 minutes, a 21-inch mercury (Hg) to 23-inch Hg vacuum (depending on the local weather barometric pressure and elevation above sea level) was developed in the flash chamber by dissipating 20-psi to 26-psi at 7-gpm to 10-gpm across the venturi injector. See Appendix A4 – Vacuum Tables as a Function of Elevation and Atmospheric Pressure. The 5,200-BTU/hr chiller and condensing loop in the flash hood was sized to maintain a constant temperature range between 64.4°F to 69.8°F, so long as the steam generation rate does not exceed the cooling capacity of the chiller and the efficiency of the condensation loop.

Once a 19-inch vacuum was reached, the brine circuit was energized by turning the rheostat controlling the power supplied to the 1.5-kW resistive heating element to a 110-V +/- 0.2 V setting. Depending on the concentration of the brine being “electro-distilled”, the amps ranged between 7-amps to 11-amps at the 110-V setting. After approximately 30-minutes, a slow boil began at 105ºF, under a 21-inch Hg to 23-inch Hg vacuum. After 35-min to 40-min, under the same vacuum, the boils turned to an aggressive at 120ºF to 130ºF as can be seen in Figure 18.—VAED flash chamber distilling 280,000-uS/cm (~179,200-mg/L TDS) brine under a vacuum.

Aggressive boil of brine at at low temperature and deep vacuum per pictures to the right.

Operating variables for picture to the left: 25-psi up | 5-psi down with 21-inch Hg vacuum | chiller at 66.4ºF

Operating variables for picture to the left: 120ºF brine temp

Power quality monitor and rheostat to control electrical load to heater

Distilled water collection trough at less than 1,630 uS/cm (~1,040 mg/L)

Figure 18.—VAED flash chamber distilling 280,000-uS/cm (~179,200-mg/L TDS) brine under a vacuum.

As the condensate evaporated from the brine, the conductivity of the brine in the flash chamber increased, which caused a temperature rise. The chiller loop was not originally sized to condense the steam as quick as it was generated because of the ohmic heating effect; therefore, the rheostat was dialed back to provide less power across the brine circuit to generate steam at a rate the 5,200-British Thermal Units per hour (BTU/hr) chiller system could condense the steam. If the power across the brine circuit was not reduced, the vacuum venturi injector system could not keep up with the positive pressure rise in the flash chamber from the steam generation and the vacuum would be lost. This resulted in elevated temperatures in the flash chamber that could melt the PVC housing. Modifications to the chilling system were made by incorporating an external car radiator and 12,000 BTU/hr air conditioner to chill the steam at the same rate it was being generated. Although this was not the most efficient design, this modification resulted in steady state operation as indicated by the vacuum remaining between 21-inch Hg and 23-inch Hg vacuum and the temperature of the brine remaining between 120ºF and 130ºF, while the flash chamber shell temperature remained below 88ºF.

When the brine evaporated to a level below the bottom of the electrode, the circuit opened and the heating element powered from the “brine circuit” de-energized, preventing overheating. The salt slurry was left in the bottom of the flash chamber as a salt seed after each run to allow the incoming, lower TDS concentrate to be increased to a conductivity that allowed efficient distillation. The only time the salt slurry was evacuated was for two sampling events where approximately 1-liter was used to fill the sample bottles.

Occasionally, the permeate water in the recirculation bath was augmented from the permeate collection tank because of evaporation.

2.3.4 Failure Mode Analysis Operation

A FMA is a critical portion of research where the limits of the treatment train are tested to reveal vulnerabilities that could play a role in further research. Lab analysis budget was exhausted by the time of the FMA; however, field conductivity, pH, temperature, ORP and flow rate was trended throughout the treatment train with the use of the E+H instrumentation.

Upon completion of the field testing in November 2021, the pilot trailer was not winterized because it was going to enter the FMA phase, where the EC + MF + RO train was going to be operated in the configuration that achieved the lowest hardness removal entering the RO system. This was intended to reveal the effect a higher foulant loading had on the RO recovery and specific flux. During the time frame between completion of the field testing and the initiation of the FMA phase, the space heater in the trailer tripped and several components of the RO pretreatment skid broke such as two of the inline pH, ORP, and temperature probes and 20-micron and 0.2-micron cartridge filter housing. The 3-stage RO system, including the high-pressure plunger pump, hoses, gauges, and RO modules were carefully tested and did not suffer freeze damage.

To continue the FMA, the pretreatment components previously mentioned needed to be replaced. A decision was made to reconfigure the 20-micron and 0.2-micron cartridge filtration in a parallel train to help decrease the hydraulic loading rate on the filters and increase the filtration surface area so more solids could be captured. The 20-micron cartridge filter was replaced with a 5-micron filter and the 0.2-micron filter was left in the lag/polishing position. The two in-line instrumentation probes were not replaced; however, the subsequent probes that remained operational were used to help draw conclusions and trends with the FMA.

To challenge the system, the RO was operated under a stressed condition of a minimal HRT of one minute, no pH adjustment, and fouled electrodes with 100% of the concentrate recycled in a proprietary manner. This required the feed pressure to increase from an average of 346 psi during the pilot to 600-psi +/- 20-psi. Initially, RO fouling was not noticed if a brief 10 gallon to 15 gallon permeate flush was implemented. To further challenge the system, the electrocoagulated and filtered blowdown water was left in the membranes twice for over three weeks with no sodium bisulfite preservative. The first time, the system was started up after a 14-gallon permeate flush, and it fully recovered. The second time, the RO was not flushed, and it still recovered to within 5% of the previous specific flux and recovery stated in table ES-1.

Throughout 2023, the system continued operating in FMA, often with months of downtime between operation both in hot and cold seasons with no SBS pickling agent left in the RO membranes or any preservatives elsewhere in the system. Only one EC blade CIP occurred on June 7, 2023, after which a much denser e-floc was noticed, indicating higher foulant removal efficacy, but not verified with lab analysis. Within approximately 730-gallons of processed blowdown, the e-floc density significantly diminished; however, the specific flux and recovery remained the same as when a denser e-floc was formed.

As of January 2024, the system continues to operate in FMA with no pretreatment other than EC at a 1-minute HRT with fouled electrode surfaces. No membrane cleanings have occurred, nor was disinfectant or caustic added to the permeate tank since the pilot began in August 2021. There was no apparent biological growth in the unit processes downstream of the EC such as the clarifier, clear cartridge filter housings, RO elements, CST, and permeate storage tank. The two parallel, 5-micron filters were rinsed and reused nine times in 2023 and the 0.2-micron filters were only rinsed and reused once in 2023; thereby, indicating a floc particle size greater than 5-microns in the clarifier supernatant. A pressure differential of 15-psi across the cartridge filters triggered a cleaning routing that consisted of spraying the filter cartridges with potable water from a garden hose, followed by a brief soak in permeate in a 5-gallon bucket.

3.0 Process Variables

3.1 How Bench Testing Results were used to Determine Optimal Pilot Configuration

The EC pretreatment to the membrane system was the workhorse of the main process train. Without the EC in place to precipitate the primary membrane foulants (i.e., hardness, silica, total suspended solids [TSS], and biological growth) the RO system would have likely irreversibly fouled causing poor recovery and excessive energy intensity. Therefore, our team performed extensive bench testing, prior to pilot testing, with 19 different tests to determine which configuration resulted in the best hardness removal efficiency as a surrogate for the best water quality feeding the RO system in the pilot. Because pH plays a significant role in hardness removal, it was also trended alongside hardness data points. Figure 3 shows the RRCC students conducting the bench tests under the supervision of the EC manufacturer and Garver's principal investigator. To align with budget, an Orbeco-Hellige MC500 Multiparameter Colorimeter was used during the bench tests for total hardness tests along with a calibrated HACH Pen Pal pH meter so multiple lab analysis did not have to be submitted to the lab.

Results of the 19 bench test configurations are presented in figure 19 with respect to total hardness removal efficiency and pH. It is important to note that the hardness reagent for the MC500 Multiparameter Lab required the pH to be between 4 to 10. The tests that showed the greatest hardness removal were between 9.7 and 10.5. Since these hardness results were ran at the maximum pH value, the accuracy of the results was questioned. Additionally, the impact of EC introducing free electrons in the water was not known by the manufacturer, now Lovibond Inc.

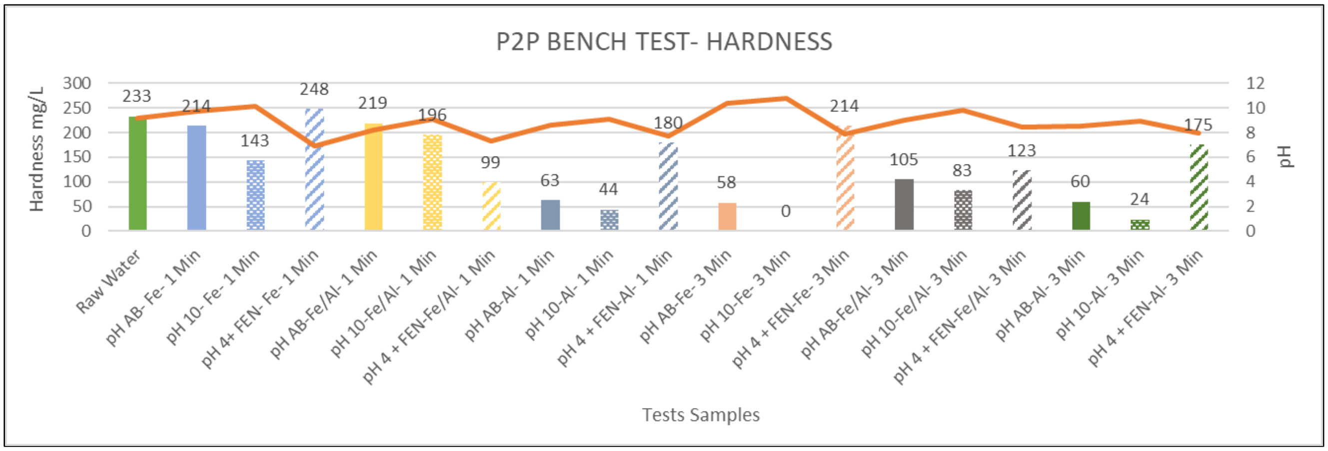


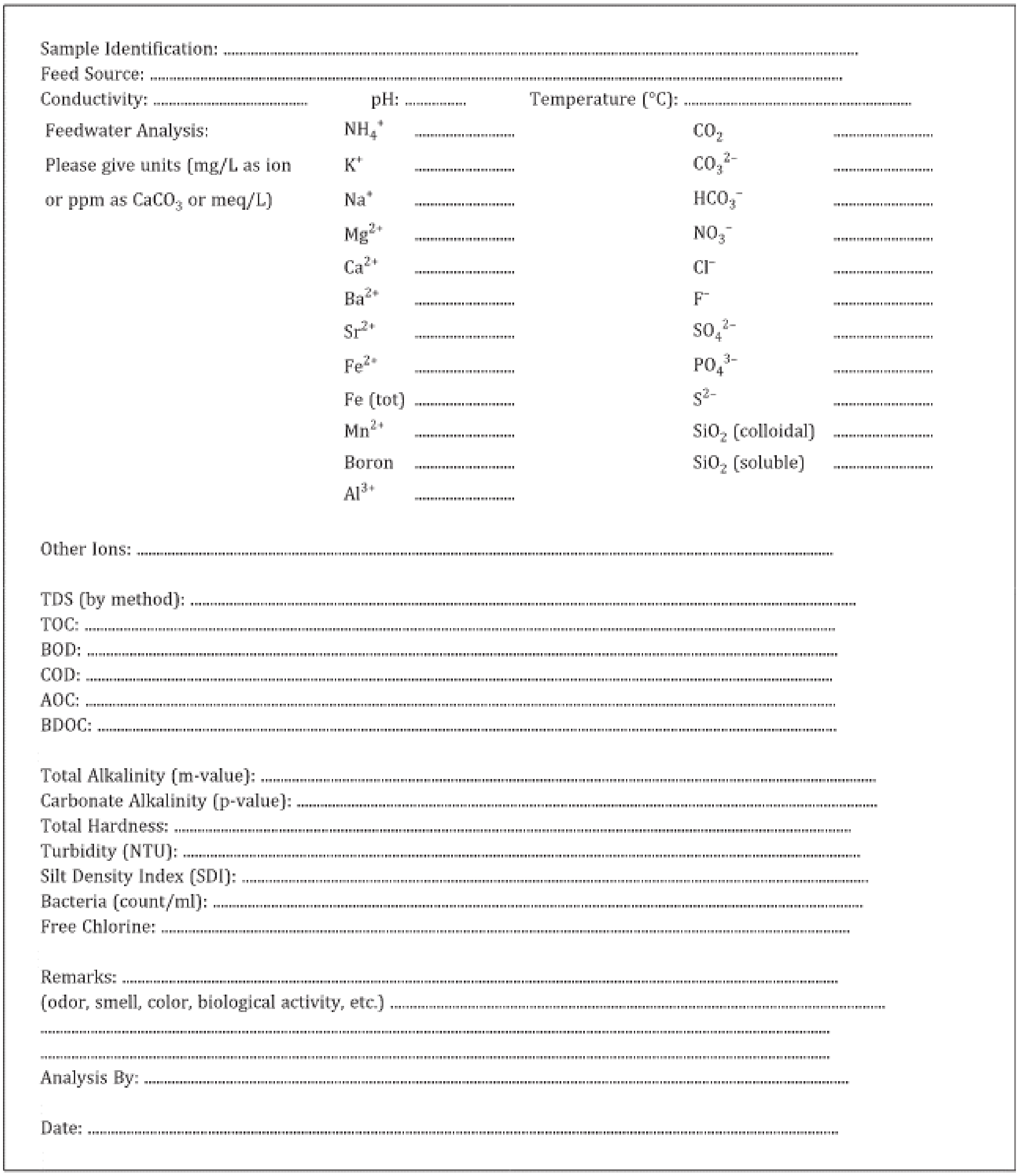
Figure 19.—Bench test results of EC on raw blowdown.

The top seven tests that had the highest hardness removal efficiency were replicated in the pilot train over the course of the 5-month pilot testing. It was decided to use a combination of iron and aluminum blades in the pilot to also address silica, under the advisement of Powell Water. The results for 23 different analyses were collected from nine different locations across the pilot train and trended in Section 3.

There was not a bench top unit for the VAED system available, so various configurations were tested during the pilot phase by adjusting operation as noted in Section 2.3.3.

3.1.1 Lab Data

Because of budget constraints, the research team had to decide whether to perform a lab analysis for all the analytes recommended by Dupont, for the DOW FilmTec™ RO membranes, at limited locations across the treatment train, or a reduced analyte list at multiple locations in the treatment train per Figure 7.—Garver's detailed process flow diagram. The decision was made to analyze a reduced list of parameters from what is recommended in Dupont' FilmTec™ Reverse Osmosis Membranes Technical Manual, Revision 8, September 2021, table 9 as seen in Figure 20.—DOW FilmTec™ RO recommended analyte list, to what was analyzed at six different locations across the main process train for the various sampling events. This allowed the team to trend treatment efficacy across the main process train to determine the contribution of each unit process in removing major constituents.



Sample Identification: ....
Feed Source: ....
Conductivity: .... pH: .... Temperature (°C): ....

Feedwater Analysis:	NH_4^+	....	CO_2	....
Please give units (mg/L as ion	K^+	....	CO_3^{2-}	....
or ppm as $CaCO_3$ or meq/L)	Na^+	....	HCO_3^-	....
	Mg^{2+}	....	NO_3^-	....
	Ca^{2+}	....	Cl^-	....
	Ba^{2+}	....	F^-	....
	Sr^{2+}	....	SO_4^{2-}	....
	Fe^{2+}	....	PO_4^{3-}	....
	Fe (tot)	....	S^{2-}	....
	Mn^{2+}	....	SiO_2 (colloidal)	....
	Boron	....	SiO_2 (soluble)	....
	Al^{3+}	....		

Other Ions: ....

TDS (by method): ....
TOC: ....
BOD: ....
COD: ....
AOC: ....
BDOC: ....

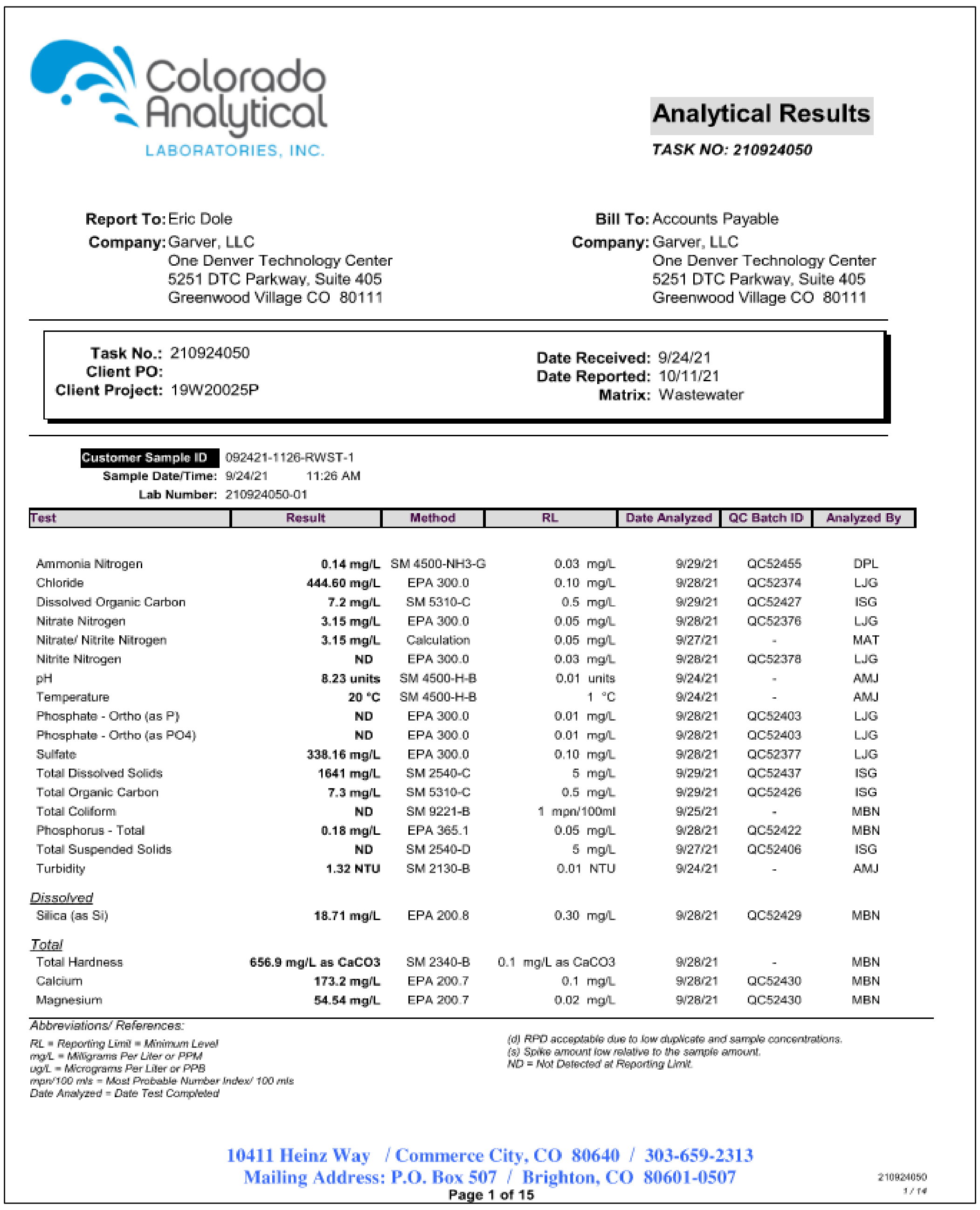
Total Alkalinity (m-value): ....
Carbonate Alkalinity (p-value): ....
Total Hardness: ....
Turbidity (NTU): ....
Silt Density Index (SDI): ....
Bacteria (count/ml): ....
Free Chlorine: ....

Remarks: ....
(odor, smell, color, biological activity, etc.) ....
....
....
Analysis By: ....

Date: ....

Figure 20.—DOW FilmTec™ RO recommended analyte list.

For a list of the reduced analytes collected and analyzed in the lab, please refer to Figure 21.—Sample lab results table for September 24, 2021. This figure represents the raw water analysis prior to caustic addition collected on September 24, 2021. These same analytes were collected at all the other sample taps in the main process train (EC/C/MF/RO) for the sample collection event that day. A total of 23 sampling routines occurred during the field testing between August 2021 and November 2021, representing seven different EC pretreatment configurations. VAED water quality sampling only occurred three times during the field testing because of the batch processing operation of the system.



Colorado Analytical
LABORATORIES, INC.

Analytical Results

TASK NO: 210924050

Report To: Eric Dole
Company: Garver, LLC
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village CO 80111

Bill To: Accounts Payable
Company: Garver, LLC
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village CO 80111

Task No.: 210924050
Client PO:
Client Project: 19W20025P

Date Received: 9/24/21
Date Reported: 10/11/21
Matrix: Wastewater

Customer Sample ID 092421-1126-RWST-1
Sample Date/Time: 9/24/21 11:26 AM
Lab Number: 210924050-01

Test	Result	Method	RL	Date Analyzed	QC Batch ID	Analyzed By
Ammonia Nitrogen	**0.14 mg/L**	SM 4500-NH3-G	0.03 mg/L	9/29/21	QC52455	DPL
Chloride	**444.60 mg/L**	EPA 300.0	0.10 mg/L	9/28/21	QC52374	LJG
Dissolved Organic Carbon	**7.2 mg/L**	SM 5310-C	0.5 mg/L	9/29/21	QC52427	ISG
Nitrate Nitrogen	**3.15 mg/L**	EPA 300.0	0.05 mg/L	9/28/21	QC52376	LJG
Nitrate/ Nitrite Nitrogen	**3.15 mg/L**	Calculation	0.05 mg/L	9/27/21	-	MAT
Nitrite Nitrogen	**ND**	EPA 300.0	0.03 mg/L	9/28/21	QC52378	LJG
pH	**8.23 units**	SM 4500-H-B	0.01 units	9/24/21	-	AMJ
Temperature	**20 °C**	SM 4500-H-B	1 °C	9/24/21	-	AMJ
Phosphate - Ortho (as P)	**ND**	EPA 300.0	0.01 mg/L	9/28/21	QC52403	LJG
Phosphate - Ortho (as PO4)	**ND**	EPA 300.0	0.01 mg/L	9/28/21	QC52403	LJG
Sulfate	**338.16 mg/L**	EPA 300.0	0.10 mg/L	9/28/21	QC52377	LJG
Total Dissolved Solids	**1641 mg/L**	SM 2540-C	5 mg/L	9/29/21	QC52437	ISG
Total Organic Carbon	**7.3 mg/L**	SM 5310-C	0.5 mg/L	9/29/21	QC52426	ISG
Total Coliform	**ND**	SM 9221-B	1 mpn/100ml	9/25/21	-	MBN
Phosphorus - Total	**0.18 mg/L**	EPA 365.1	0.05 mg/L	9/28/21	QC52422	MBN
Total Suspended Solids	**ND**	SM 2540-D	5 mg/L	9/27/21	QC52406	ISG
Turbidity	**1.32 NTU**	SM 2130-B	0.01 NTU	9/24/21	-	AMJ
Dissolved						
Silica (as Si)	**18.71 mg/L**	EPA 200.8	0.30 mg/L	9/28/21	QC52429	MBN
Total						
Total Hardness	**656.9 mg/L as CaCO3**	SM 2340-B	0.1 mg/L as CaCO3	9/28/21	-	MBN
Calcium	**173.2 mg/L**	EPA 200.7	0.1 mg/L	9/28/21	QC52430	MBN
Magnesium	**54.54 mg/L**	EPA 200.7	0.02 mg/L	9/28/21	QC52430	MBN

Abbreviations/ References:

RL = Reporting Limit = Minimum Level
mg/L = Milligrams Per Liter or PPM
ug/L = Micrograms Per Liter or PPB
mpn/100 mls = Most Probable Number Index/ 100 mls
Date Analyzed = Date Test Completed

(d) RPD acceptable due to low duplicate and sample concentrations.
(s) Spike amount low relative to the sample amount.
ND = Not Detected at Reporting Limit.

10411 Heinz Way / Commerce City, CO 80640 / 303-659-2313
Mailing Address: P.O. Box 507 / Brighton, CO 80601-0507
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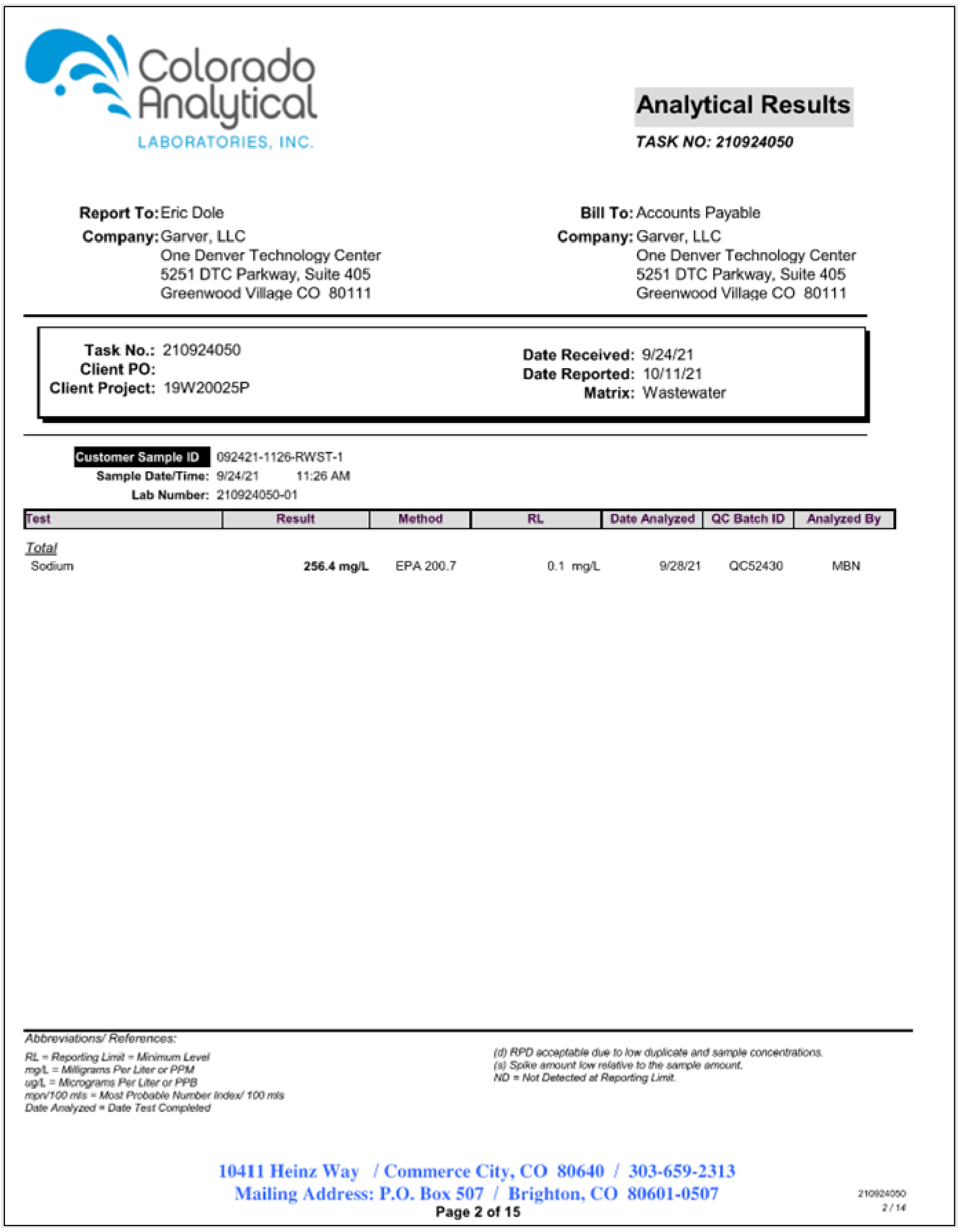
1 / 14



Colorado Analytical
LABORATORIES, INC.

Analytical Results

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Date Received: 9/24/21
Date Reported: 10/11/21
Matrix: Wastewater

Customer Sample ID 092421-1126-RWST-1
Sample Date/Time: 9/24/21 11:26 AM
Lab Number: 210924050-01

Test	Result	Method	RL	Date Analyzed	QC Batch ID	Analyzed By
Total						
Sodium	**256.4 mg/L**	EPA 200.7	0.1 mg/L	9/28/21	QC52430	MBN

Abbreviations/ References:

RL = Reporting Limit = Minimum Level
mg/L = Milligrams Per Liter or PPM
ug/L = Micrograms Per Liter or PPB
mpn/100 mls = Most Probable Number Index/ 100 mls
Date Analyzed = Date Test Completed

(d) RPD acceptable due to low duplicate and sample concentrations.
(s) Spike amount low relative to the sample amount.
ND = Not Detected at Reporting Limit.

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Figure 21.—Sample lab results table for September 24, 2021.

For ease of reference, figure 7 is presented again below along with a description of each sample tap location.

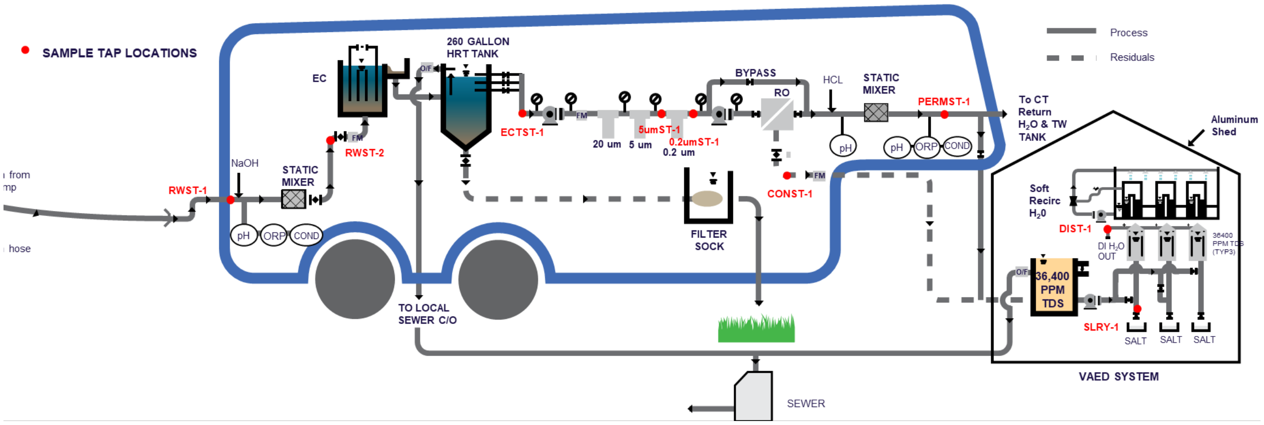


Figure 22.—Garver's detailed process flow diagram.

Sample Tap Description:

- **RWST-1:** Raw water sample tap located prior to caustic addition.
- **RWST-2:** Raw water sample tap located after caustic addition and two static mixers, but before entering the bottom of the EC system. **NOTE:** Depending on which seven testing configurations were implemented, this sample tap, and all of the ones below, may be influence by caustic addition.
- **ECST-1:** EC treated water sample tap located on the MF boost pump discharge that represents the electrocoagulated, clarified supernatant downstream of the EC unit.
- **5umST-1:** EC treated, clarified, and filtered water sample tap downstream of the 20-micron cartridge filter (during pilot) and the 5-micron (during FMA) cartridge filter.
- **0.2umST-1:** EC treated, clarified, and filtered water sample tap downstream of the 20-micron cartridge filter, the 5-micron cartridge filter, and the 0.2-micron cartridge filter. **NOTE**: For the GAC configuration, which was only one sample set because of high TSS from sloughing, the 20-micron filter was replaced with the 5-micron filter, the 5-micron filter was replaced with the 0.2-micron filter and the GAC cartridge went in the 0.2-micron cartridge.
- **PERMST-1:** Combined RO permeate sample tap from all three stages. There was no post treatment needed for the purpose of this research.
- **CONST-1:** Concentrate sample tap located downstream of the backpressure valve.

- **DIST-1:** Distillate sample tap for the VAED system.
- **SLRY-1:** Salt slurry sample tap of the leftover seeded concentrate for the VAED system.

In the following sections, there are several graphs depicting the removal efficacy of target analytes and other pertinent field operating trends. An interactive lab analysis Excel spreadsheet will be submitted with this report that allows the Reclamation staff to graph and trend various aspects of the water quality analysis.

To see a quote from Colorado Analytical Laboratory that reveals the water quality analytes sampled across the process train and the analytical method used please refer to appendix-A5.

3.1.2 Instrument Data

Special thanks to our industry partners, Endress+Hauser and Rockwell Automation, for providing the various instrumentation identified below pro-bono:

- **Rockwell Automation Power quality monitors** – Four, PowerMonitor 500 (catalog number 1420-V1P-ENT) power quality monitors were installed at the various electrical feed locations to sub-meter the power (kW) and consumption (kWh) for each unit process:
    1. Main trailer feed (includes all trailer electrical load like EC, membrane skid, lights, AC, and miscellaneous outlets).
    2. EC unit.
    3. Membrane skid (includes 1/3-HP prefilter boost pump and 1-HP RO feed pump).
    4. VAED unit.
- **Endress+Hauser In-line monitoring cells for pH (No.), temperature (ºF), ORP (mV) and conductivity (mS/cm)** – Eight CPS16D and eight CLS82D Model Sensors inside CPA250 Model Flow Assemblies were located at the various location noted below to monitor the treatment efficacy in a real-time environment:
    1. Immediately after **RWST-1** and before **RWST-2** sample tap to trend the raw water prior to caustic addition.
    2. Downstream of **RWST-2**, just before entering the bottom of the EC unit, to allow ample mixing of caustic prior to flowing through the cell.
    3. After the 20-micron cartridge filter, immediately upstream of **20umST-1** sample tap.
    4. After the 5-micron cartridge filter, immediately upstream of **5umST-1** sample tap.

5. After the 0.2-micron cartridge filter, immediately upstream of **0.2umST-1** sample tap that resembles the feed water quality entering the RO.

6. On the combined permeate, immediately before **PERMST-1** sample tap.

7. On the concentrate, immediately before **CONST-1** sample tap.

8. A spare cell was set aside to act as a tester when calibrating pH, ORP, temperature and conductivity meters or verifying water quality from the salt slurry and distillate for the VAED system that was operated in a batch mode.

- E+H Pico Mag magnetic flow meters – Seven, DMA15 Model Pico Mag meters trended real-time flow (gpm), totalized flow (kgal), temperature (°F) and conductivity (mS/cm) at the various locations noted below:

  1. Immediately before **RWST-1** sample tap to trend the raw water flow rate entering the EC to ensure accurate HRT in the EC chamber depending on the testing configuration.

  2. Immediately before the 20-micron cartridge filter to represent the flow entering the membrane skid.

  3. On the 1st Stage permeate.

  4. On the 2nd Stage permeate.

  5. On the 3rd Stage permeate.

  6. After the confluence of all three stages to represent the combined permeate water quality.

  7. On the concentrate, immediately before **CONST-1** sample tap, downstream of backpressure valve.

- E+H Data logger – One, RSG45 Model data logger compiled all the 4-20 mA data from the various E+H instrumentation and trended the values in a real-time environment with graphical outputs on the display of the logger. The data logger was located in the cabinet above the membrane skid.

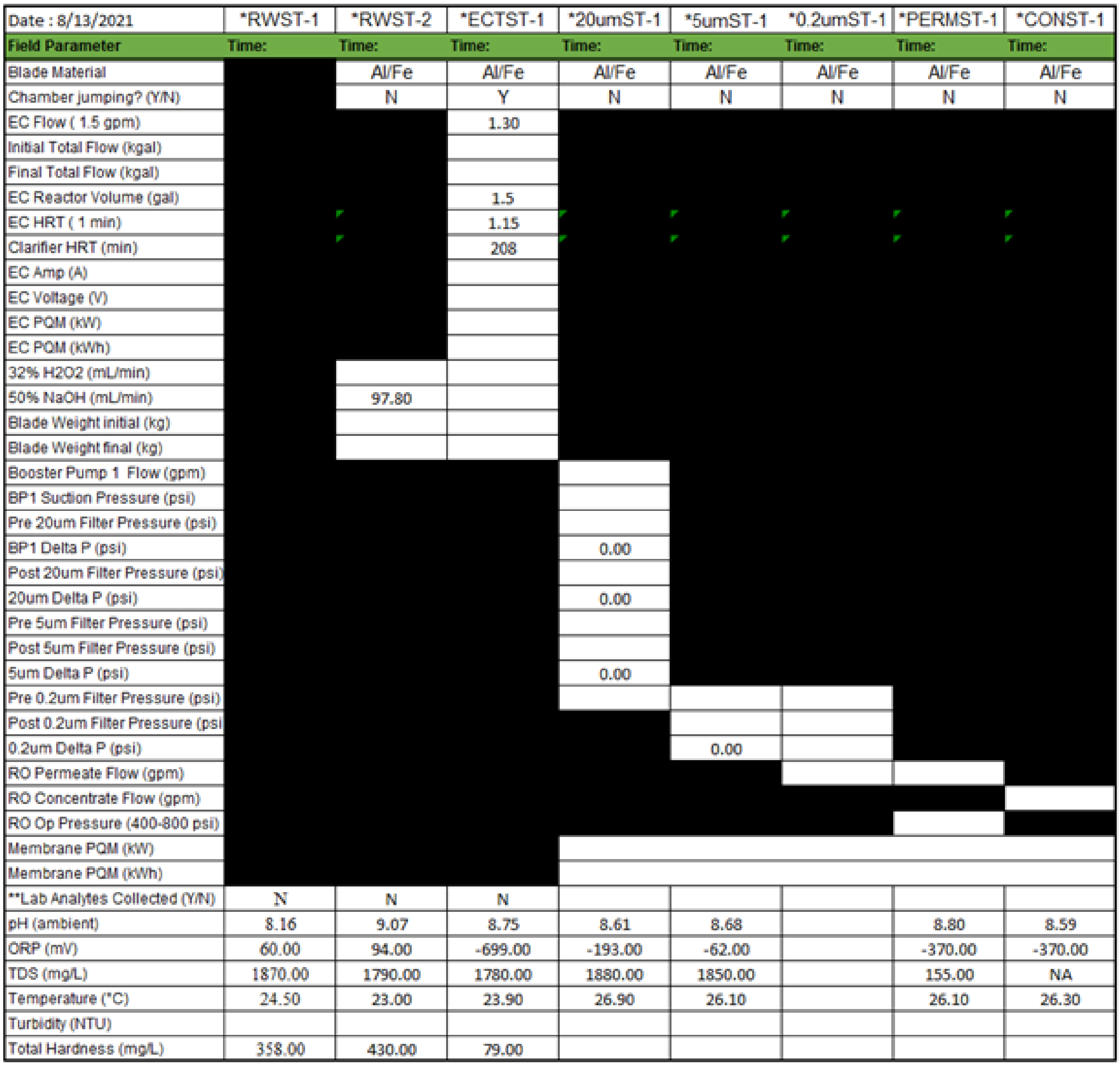
Lyons Filter graciously donated their field technician for three consecutive days to assist and train our team on the installation and operation of all the various instrumentation throughout the process train.

3.1.3 Sample Collection Procedures

The sample collection routine required significant coordination and adhered to meticulous procedures to ensure accurate data. Prior to each grab sample routine, the operator annotated each bottle set according to the date, time, and sample tap location. The bottle sets were strategically placed on the adjacent countertop in the trailer behind a piece of tape that indicated the sample tap location. In addition, each sample tap had a 250-mL glass beaker to be filled at the beginning of each grab sample event. The water collected in this sample was used to verify field data like pH, ORP, temperature, and conductivity and correlated to the sample routine time stamp that was submitted to a lab for analysis. Our research team felt this was the best way to accurately coordinate field water quality metrics to each grab sample versus trying to sift through substantial amounts of data that was trended with the data logger.

If two operators were present at the time of a grab sample collection, one filled the bottle sets, while the other recorded the field data. If only one operator was present during a grab sample event, field data was collected after bottle set was filled and during labeling. Regardless of the number of operators present during a sampling event, the sample sets were filled in the same chronological order that the water flowed through the process train, one at a time. For example, the bottles were filled in the following order: RWST-1, RWST-2 (if pH adjusted), ECST-1, 5umST-1, 0.2umST-1, GACST-1 (for only one sample set), PERMST-1, CONST-1. Care was taken not to open a sample tap all the way and affect the process train flow rate and correlating pumps.

The following table shown in Figure 23.—Sample field data collection log was populated with field data during each grab sample routine.



Date : 8/13/2021	*RWST-1	*RWST-2	*ECTST-1	*20umST-1	*5umST-1	*0.2umST-1	*PERMST-1	*CONST-1
Field Parameter	Time:	Time:	Time:	Time:	Time:	Time:	Time:	Time:
Blade Material		Al/Fe	Al/Fe	Al/Fe	Al/Fe	Al/Fe	Al/Fe	Al/Fe
Chamber jumping? (Y/N)		N	Y	N	N	N	N	N
EC Flow (1.5 gpm)			1.30					
Initial Total Flow (kgal)								
Final Total Flow (kgal)								
EC Reactor Volume (gal)			1.5					
EC HRT (1 min)			1.15					
Clarifier HRT (min)			208					
EC Amp (A)								
EC Voltage (V)								
EC PQM (kW)								
EC PQM (kWh)								
32% H2O2 (mL/min)								
50% NaOH (mL/min)		97.80						
Blade Weight initial (kg)								
Blade Weight final (kg)								
Booster Pump 1 Flow (gpm)								
BP1 Suction Pressure (psi)								
Pre 20um Filter Pressure (psi)								
BP1 Delta P (psi)				0.00				
Post 20um Filter Pressure (psi)								
20um Delta P (psi)				0.00				
Pre 5um Filter Pressure (psi)								
Post 5um Filter Pressure (psi)								
5um Delta P (psi)				0.00				
Pre 0.2um Filter Pressure (psi)								
Post 0.2um Filter Pressure (psi								
0.2um Delta P (psi)					0.00			
RO Permeate Flow (gpm)								
RO Concentrate Flow (gpm)								
RO Op Pressure (400-800 psi)								
Membrane PQM (kW)								
Membrane PQM (kWh)								
**Lab Analytes Collected (Y/N)	N	N	N					
pH (ambient)	8.16	9.07	8.75	8.61	8.68		8.80	8.59
ORP (mV)	60.00	94.00	-699.00	-193.00	-62.00		-370.00	-370.00
TDS (mg/L)	1870.00	1790.00	1780.00	1880.00	1850.00		155.00	NA
Temperature (°C)	24.50	23.00	23.90	26.90	26.10		26.10	26.30
Turbidity (NTU)								
Total Hardness (mg/L)	358.00	430.00	79.00					

Figure 23.—Sample field data collection log.

Prior to collecting samples, the operator allowed the system to run for one hour to ensure steady state operations. While working to achieve a steady state all field instruments were verified for accuracy and all glassware was inspected for cleanliness.

Since the VAED system operated as a batch process, because it required a higher conductivity brine than the current pilot skid could provide, only three grab samples were sent to the lab for analysis. All the same analytes were tested as indicated in Figure 31.—VAED lab analysis for the three grab sample sets vs the concentrate concentrations. The three grab sample collection dates were October 20, 2021, November 17, 2021, and November 23, 2021.

Constantly distilling the brine in the flash chamber, and not blowing down the salt slurry after seven of the eight gallons of the concentrate/salt seed batch was flashed off and collected as distillate, resulted in a moderate increase in conductivity. When the operator felt the conductivity was high enough to send to the lab, the operator allowed the system to cool to ambient temperature (65 Fahrenheit (F) to 78°F) before 1-L of the 3-L salt slurry remaining in bottom of the flash chamber was evacuated from the flash vessel into the 10-gallon high density polyethylene (HDPE) collection trough below. The operator mixed the salt slurry in the trough with a PVC pipe to ensure a homogeneous mixture was used to fill the bottle sets. Once mixed, the salt slurry was carefully transferred into the respective bottle sets and labeled. The distillate was evacuated from the distillate collection trough and filled the respective bottle sets directly from the sample tap. Distillate samples were collected by draining the distillate trough into the sample bottles after collecting the slurry samples.

3.2 Independent Variables

3.2.1 Controlled

Controlled variables for the main process train (EC/CLAR/MF/RO), such as flow rate, HRT, pH, chamber electrode configuration, RO backpressure/percent recovery and polarity reversal frequency all have an impact in the performance of the system. VAED controlled variables consisted of electrode configuration, brine concentration in the flash chamber, ohmic heating intensity, level of vacuum, chiller BTU setting, and condensation hood configuration. Each of these controlled variables, and the impact they had on the process train efficiency will be discussed in further detail in Section 5.

3.2.2 Uncontrolled

Uncontrolled variables also impacted the treatment efficiency of the pilot trailer and needed to be accounted for during operation of the field testing. Items such as frequency of blowdown, seasonal impacts, chemical treatment of recirc water in the cooling tower loop and raw water characteristics of blowdown all impacted the treatment. More detail of the impacts of these uncontrolled, independent variables will also be discussed in Section 5.

3.3 Dependent Variables

As revealed in Section 4, dependent variables such as the water quality results, percent recovery, specific flux, temperature, and energy intensity varied depending on the configuration of the system. The FMA tests revealed positive trends not anticipated in the beginning of the research and will be discussed further in Section 5.

4.0 Results and Discussion

4.1 Water Quality Analysis

A significant effort was made to develop an interactive graphing tool that allowed trending of constituents from the lab analysis at nine separate locations throughout the treatment train for all 23 sampling events representing seven different tests. This electronic tool will be submitted to the Reclamation with this report, so they can analyze other trends at their discretion. The following sections reveal the data our team feel is pertinent to the goals of this research.

4.1.1 Data Table

There were seven shortlisted scenarios developed from the bench tests where several treatment configurations were tested as detailed in Table 2.—EC pilot testing configurations and average field results.

Table 2.—EC pilot testing configurations and average field results

Skid Config.	EC/CLAR/20um/5um/0.2um/3stg RO						
Test #	*1	2	3	4	5	6	7
Testing Constraints	10.3 pH/CON to EC/1 min	AMB pH/ No CON Recovery/1 min	AMB pH/CON to EC/1 min	10.3 pH/CON to Clarifier/1 min	10.3 pH/No CON to EC/1 min	10.3 pH/CON to EC/3min	10.3 pH/No CON Recovery/3min
# of Samples Analyzed	8	4	4	2	2	1	2
Average Flow to EC	1.6	1.5	1.5	1.5	1.45	0.52	0.5
Average Feed Pressure (psi)	359	335	325	370	320	300	410
Average Permeate BP (psi)	40	32	35	45	33	27	48
Average Permeate Flow (gpm)	0.86	0.805	0.84	0.875	0.78	0.63	0.80
Average RO Feed TDS (mg/L)	1666	1584	2000	1575	1891	1527	1471
Average Permeate TDS (mg/L)	17.5	6.2	6.8	5.0	8.5	11.0	18.5
Average Specific Flux (gfd/psi)	0.047	0.047	0.052	0.047	0.048	0.041	0.038
Average Perm Flux (gfd)	14.2	13.3	13.8	14.5	12.8	10.4	13.3
Average Concentrate Flow (gpm)	0.81	0.87	0.85	0.80	0.91	1.05	0.87
Average % Recovery 3-stg RO	52%	48%	50%	52%	46%	38%	48%
Average % Recovery Overall @ 400 ppm TDS w/ Blend	58%	55%	55%	59%	52%	45%	56%

*Only one sample was tested where filtration scheme was changed to 0.5um/0.2um/GAC. Once labs showed TSS increase for RO feed from this configuration the GAC was permanently removed from the train. Sloughing of GAC could have cause dammage to membrane and GAC did not provide any additional TOC removal.

For locations of all six separate locations across the process train please refer to figure 22. For the full list of analytes sent to the lab for each sample set please refer to appendix A-3.

Below is a brief description of each of the seven testing scenarios described in table 4-1:

- **Test No. 1 – pH 9.7–– 10.3/Concentrate Recovery to EC/1-min HRT**
  An average blowdown flow rate of 1.6-gpm was sent to the process train, while the pH increased from approximately 8.6 to between 9.7 and 10.3. After traveling through 16 feet of ½ inch schedule 40 PVC pipe and two 8-inch long by ½ inch static mixers, flow entered the bottom of the EC unit. The EC unit was configured as one chamber without electrode jumping, which resulted in 3 to 6 amps at 100 volts. The pH adjusted blowdown flowed upwards through the energized EC unit, where an instantaneous floc was formed and allowed to settle in the clarifier immediately downstream. The clear supernatant was pumped from 4 inches below the water surface with the use of the floating weir. The variable frequency drive boost pump fed the membrane skid in the configuration listed in table 4-1 and would ramp up/down to maintain a set pressure range of 30 psi to 40 psi to the RO feed pump, depending on head loss across filters. The RO feed pump increased the pressure to an average of 359 si, depending on the throttle position of the concentrate needle valve. Approximately 10% to 20% of the concentrate flow was returned to the head of the plant prior to pH adjustment to increase the incoming conductivity and decrease the amount of concentrate sent to the VAED and to increase blowdown TDS.

- **Test No. 2 – Ambient pH 8.6/No Concentrate Recovery/1-min HRT**
  An average blowdown flow rate of 1.5-gpm was sent to the process train without pH adjustment at an ambient pH of 8.6. The same flow path was taken as described in Test No. 1, except all the concentrate was sent to VAED system with no recirculation back to head of treatment. The RO feed pump increased the pressure to an average of 335 psi.

- **Test No. 3 – Ambient pH 8.6/Concentrate Recovery to EC/1-min HRT**
  An average blowdown flow rate of 1.5-gpm was sent to the process train without pH adjustment at an ambient pH of 8.6. The same flow path was taken as described in Test No. 2, except 10% to 20% of the concentrate flow was recirculated back to head of treatment. The RO feed pump increased the pressure to an average of 325-psi.

- **Test No. 4 – pH 9.7–– 10.3/Concentrate Recovery to Clarifier/1-min HRT**
  An average blowdown flow rate of 1.5-gpm was sent to the process train, while increasing the ambient pH from approximately 8.6 to between 9.7 and 10.3 as previously noted. The same flow path was taken as described in Test No. 3, except 10 to 20% of the concentrate flow was recirculated back to the clarifier after the EC unit. The RO feed pump increased the pressure to an average of 370 psi.

- **Test No.5 – pH 9.7–- 10.3/No Concentrate Recovery/1-min HRT**
  An average blowdown flow rate of 1.45-gpm was sent to the process train, while increasing the ambient pH from approximately 8.6 to between 9.7 and 10.3 as previously noted. The same flow path was taken as described in Test No.4, except all the concentrate was sent to VAED system with no recirculation back to head of treatment. The RO feed pump increased the pressure to an average of 320-psi.

- **Test No.6 – pH 9.7–- 10.3/Concentrate Recovery to EC/3-min HRT**
  An average blowdown flow rate of 0.52-gpm was sent to the process train, while increasing the ambient pH from approximately 8.6 to between 9.7 and 10.3 as previously noted. The same flow path was taken as described in Test No.5, except 10% to 20% of the concentrate flow was recirculated back to the clarifier after the EC unit. The RO feed pump increased the pressure to an average of 300-psi. As previously noted, in this arrangement the clarifier was run in batch mode. The contents of the clarifier were processed through the system through the floating weir to a level just over the solids blanket prior to filling the 270-gallon clarifier with the 3-min HRT electrocoagulated water. This was done to purge the contents of the clarifier with processed water remaining from the previous 1-min HRT runs but leaving the solids blanket. Once full, the EC system would continue to process raw water at the 0.5 gpm flow rate, while the membrane system would process water at the 1.6-gpm flow rate. This caused the clarifier to drain and not act as a true clarifier at steady state as was the case with the 1-min HRT.

- **Test No.7 – pH 9.7–- 10.3/No Concentrate Recovery/3-min HRT**
  An average blowdown flow rate of 0.5-gpm was sent to the process train, while increasing the ambient pH from approximately 8.6 to between 9.7 and 10.3 as previously noted. This reduced flow rate required a batch operation as previously described in Test No.6. The same flow path was taken as described in Test No.6, except all the concentrate was sent to VAED system with no recirculation back to head of treatment. The RO feed pump increased the pressure to an average of 410-psi, which was the highest pressure seen during these tests. The same batch operation was incorporated in this test as in Test No.6.

In all seven of the tests, backpressure was induced on the first and second stage permeate to increase crossflow velocity in the absence of multiple parallel arrays. A brief permeate flush of the RO system was initiated at 0% recovery prior to each run until the conductivity of the concentrate was double that of the permeate used to scour/rinse the membrane surface of foulants, as indicated by the in-line conductivity analyzers. When the concentrate conductivity dropped to within 20% of the permeate feed flush conductivity, the membranes were considered flushed/cleaned and the specific flux was recovered without the need for traditional CIP chemicals. The volume of permeate rinse ranged from 10 to 25-gal depending on the test configuration and runtime. The max recovery was determined by the position of the concentrate valve and what the operator felt resulted in a max pressure without fouling of the membrane, or a consistent rise in pressure at a set valve position.

There were three different electrode configurations and two different cooling configurations with the VAED system. Because of the confidential nature of the patent and its stages of development, only a high-level description will be provided regarding the results of the various configurations. In the following sections.

4.1.2 Analysis/Observations –- Feedwater

Two major observations were noticed in the pilot trailer feedwater. They are discussed in detail below.

4.1.2.1 Cation/Anion Balance of Feed Water Lab Data

Review of lab data obtained from analysis of submitted samples revealed the following patterns:

- Measured TDS > estimated TDS
    - Greater than 85% of the lab measurements of TDS were higher than TDS estimated from summing up the lab measurements of major ions and silica
- Cation/Anion balance > 1
    - This occurred in >75% of the cases
- Calcium concentrations after NaOH addition < calcium concentrations prior to addition
    - This occurred in the majority of cases

"The sample containers for cation measurements have acid in them to facilitate Inductively Coupled Plasma (ICP) measurement metal concentrations. As a result, cation concentrations tend to be high as they may redissolve some colloid/solids.". This explains the TDS and cation/anion balance observations.

In the case of calcium measurements, the explanation was that not enough acid was present to solubilize all the colloidal/solid $CaCO_3$ because of the addition of caustic. When this occurs, the lab filters the sample so as not to clog the ICP equipment ports.

The importance of the observations and explanations is that care must be taken to match analytical measurements taken with the results sought. In the project the sought results were for dissolved concentrations and the data obtained did not permit a more detailed analysis of what fraction of hardness removal was because of pH change and what fraction was because of the EC treatment.

4.1.2.2 Blowdown Restriction Impact on Raw Water Quality

After a month of runtime, an extra Pico magmeter (sized for 0.2-gpm to 20-gpm flow rate) was available to estimate the actual flow and TDS mass loading to the Blowdown Storage Tank. The research team initially thought that this would be a good method to not only quantify the daily volume of saline blowdown, but also record the conductivity (or TDS) of the blowdown so an approximation of TDS mass loading to the sewer could be estimated. The Pico Mag uses Faraday's Law, which relies on the conductivity of the water, to measure flow; therefore, it reports an estimation of the water conductivity along with flow, total flow, and temperature.

As can be seen in Figure 24.—RWST-1 and 0.2UMST-1 and GACST-1: total hardness vs. total dissolved solids, the total hardness as $CaCO_3$ was approximately 650-mg/L. The Pico Mag was installed on the blowdown line on 9/27/2022. During the last week of October, our team noticed the RWST-1 conductivity values from the in-line conductivity sensor were steadily increasing. Because the RWST-1 sample tap was before the concentrate recirculation location, the increase in conductivity was not attributed to the test being performed at the time. Upon further investigation, our team realized the ¼" square opening in the center of the Pico Mag was causing a severe restriction on the blowdown line acting like a throttled valve, as opposed the ½" diameter PVC fill line it previous flowed through. This resulted in less water leaving the recirculation loop as blowdown and preventing less potable water to dilute the water quality.

The Pico Mag was immediately removed and the raw water (RW) hardness, the solid red line, and TDS, the dashed red line, from the labs showed immediate reduction. This was expected since the Pico Mag restriction was no longer preventing the correct volume of high hardness, high TDS blowdown from leaving the system, so fresh potable water can be introduced to the cooling tower water and dilute hardness and TDS concentrations.

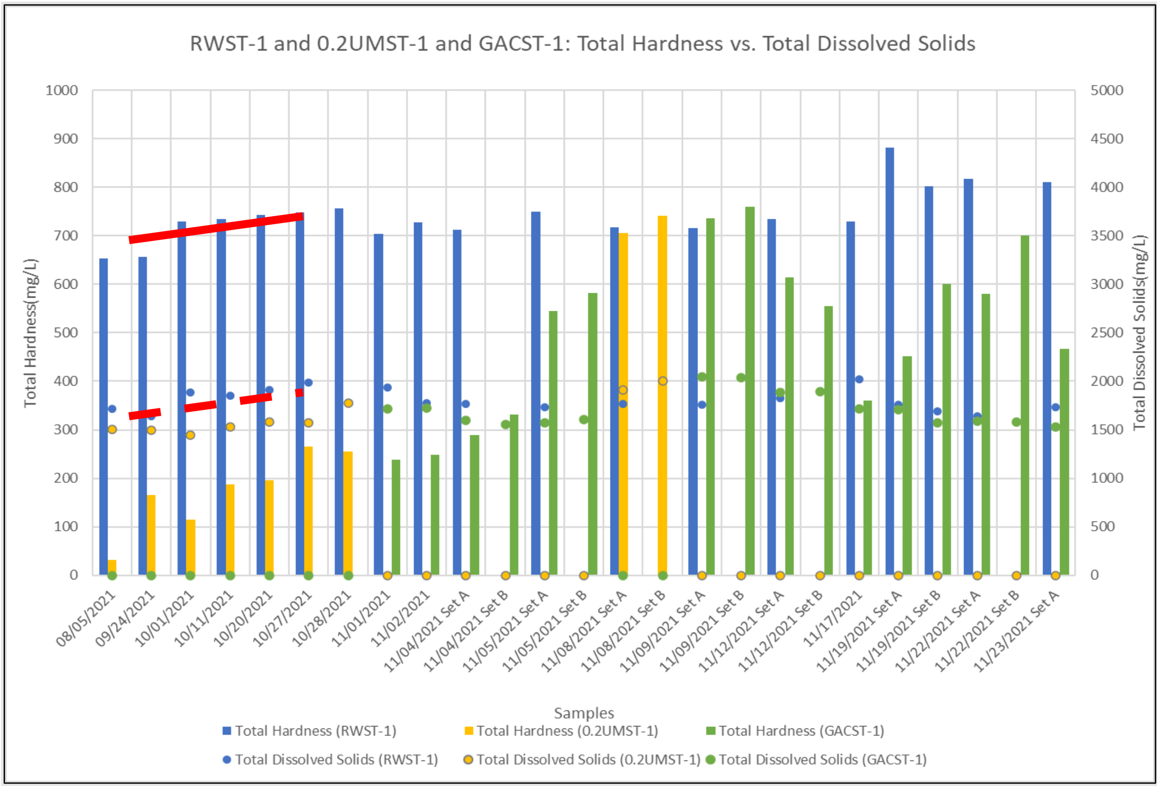
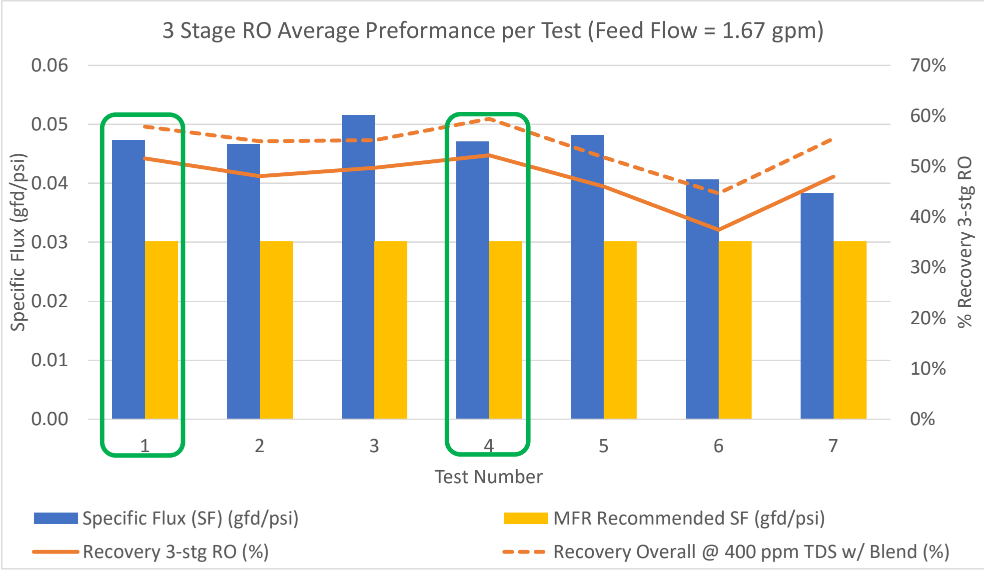


Figure 24.—RWST-1 and 0.2UMST-1 and GACST-1: total hardness vs. total dissolved solids.

4.2 Notable Trends Discussion

Over 22 different constituents were analyzed (per Figure 21.—Sample lab results table for 9/24/21, at seven different locations in the main process train on 23 grab sample events representing seven different testing configurations. Only three grab sample events occurred with the VAED system, because of the batch nature of the process and the need to have high conductivity water for it to work effectively. The VAED system only had two sampling locations that represented the distillate and the salt slurry left after a flash occurred.

Because of the exorbitant combination of trends that can be analyzed for the main process train (EC/CLAR/MF/RO), our team decided to focus on which runs had the highest recovery, specific flux, and TDS removal as an indication of treatment efficiency. Refer to Figure 25.—3-Stage RO performance per test.



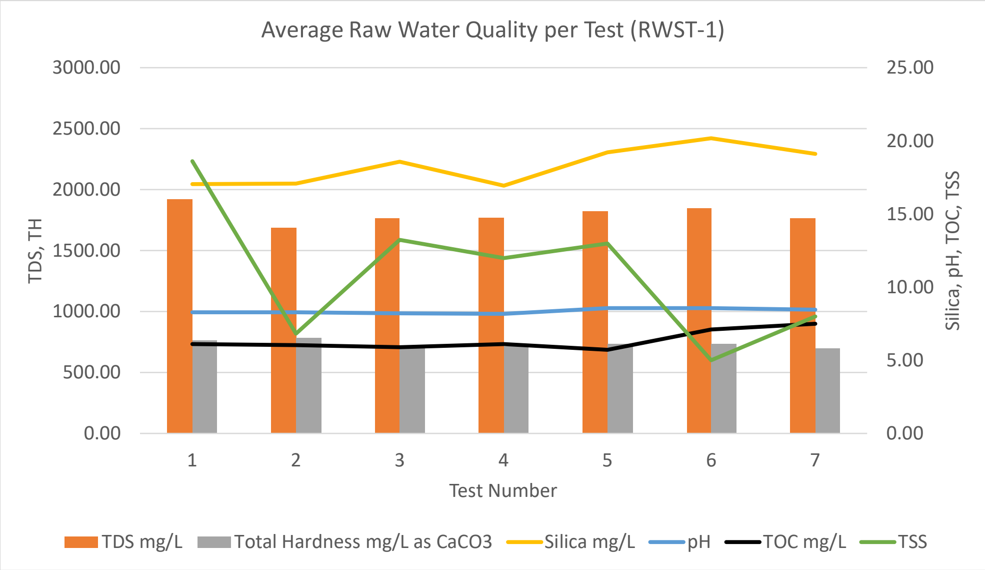
Test #	Testing Constraints	# of Samples Analyzed
1	10.3 pH/CON to EC/1 min	8
2	AMB pH/ No CON Recovery/1 min	4
3	AMB pH/CON to EC/1 min	4
4	10.3 pH/CON to Clarifier/1 min	2
5	10.3 pH/No CON to EC/1 min	2
6	10.3 pH/CON to EC/3min	1
7	10.3 pH/No CON Recovery/3min	2

Figure 25.—3-Stage RO performance per test.

The two green boxes in the graph represent the top two performing tests; however, Test No.1 had eight samples whereas Test No.4 only had two. Because Test No.1 has a greater sample size, it is recommended that the trend in Test No.1 be more reliable than the trends in Test No.4 with respect to specific flux and recovery.

The greater the specific flux, the greater the permeability, while still removing greater than 99.5% of the salt. An important trend to note, is the specific fluxes realized in the field during the pilot as identified in figure 25 are roughly 70% more than what the manufacturer specifies for the seawater membrane.

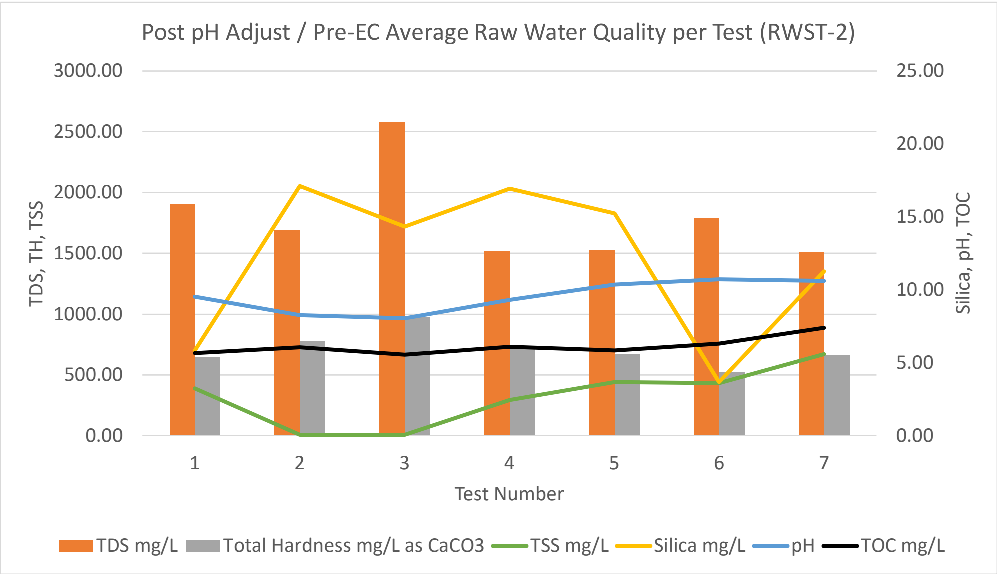
For each configuration, the research team shortlisted the 22 analytes to what were the most prudent ones to trend. They are as follows: TDS, total hardness, silica, pH, TOC, and TSS. The graphs below represent the lab values for each of the previously mentioned analytes at the six separate locations in the process train as describe in detail in Section 3.1.1 The values on the primary and secondary y-axis were kept the same to make the trends more apparent, except for the concentrate graph as annotated in the secondary y-axis as red.



Test #	Testing Constraints	# of Samples Analyzed
1	10.3 pH/CON to EC/1 min	8
2	AMB pH/ No CON Recovery/1 min	4
3	AMB pH/CON to EC/1 min	4
4	10.3 pH/CON to Clarifier/1 min	2
5	10.3 pH/No CON to EC/1 min	2
6	10.3 pH/CON to EC/3min	1
7	10.3 pH/No CON Recovery/3min	2

Figure 26.—Average Raw Water Quality per Test (RWST-1).

The raw water blowdown TDS, pH, TOC, silica, and hardness concentrations remained fairly consistent throughout all seven tests, whereas TSS varied between 5-mg/L and 18-mg/L. This wider range of average TSS concentrations may be a factor of the local weather as airborne particulates can be introduced to the cooling tower recirc loop during windy events. The slight increase in hardness noticed from the blowdown flow rate restriction previously discussed was not as apparent when the values were averaged over the time frame.



Test #	Testing Constraints	# of Samples Analyzed
1	10.3 pH/CON to EC/1 min	8
2	AMB pH/ No CON Recovery/1 min	4
3	AMB pH/CON to EC/1 min	4
4	10.3 pH/CON to Clarifier/1 min	2
5	10.3 pH/No CON to EC/1 min	2
6	10.3 pH/CON to EC/3min	1
7	10.3 pH/No CON Recovery/3min	2

Figure 27.—Post pH adjust/pre-EC average raw water quality per test (RWST-2).

Results presented in this section represent the impact caustic addition had on the water quality as this sample tap was located before the EC system. The silica increase from the caustic addition, prior to the EC, was expected as silica solubility increases to 800-mg/L at a pH of 10 as discussed in Section 4.2.1 for a more detailed explanation of this phenomenon.

RWST-1 to RWST-2 Trends

TDS: Test No.1 did not see any change in TDS likely because any reduction in the hardness component of TDS value was offset by the recirculated concentrate. Test No.2 also did not see a reduction in TDS because there was not a RWST-2 sample collected since it was the same as RWST-1. Test No.3 had a 46% increase in TDS likely because marginal hardness removal occurred with the 1-minute HRT and no pH increase, combined with the recirculated concentrate, resulted in elevated TDS concentrations at the RWST-2 tap. Tests No.4 and No.5 had a 14% and 16% reduction in TDS, respectively, because the increased pH likely removed the hardness component of TDS. The concentrate recycle did not appear to make a difference in these two tests. Test No.6 did not experience a change in TDS, whereas Test No.7 had a 14% reduction in TDS, and it did not recycle the concentrate.

Total Hardness: Test No.1 had a 15% reduction in total hardness likely because of the caustic addition. Test No.2 also did not see a change in total hardness because there was not a RWST-2 sample collected since it was the same as RWST-1. Test No.3 had a 36% increase in total hardness likely because marginal hardness removal occurred with the 1-minute HRT and no pH increase, combined with the recirculated concentrate, resulted in elevated total hardness concentrations at the RWST-2 tap. Tests No.4 and No.5 did not have a significant change in total hardness. The concentrate recycle did not appear to make a difference in these two tests. Test No.6 had a

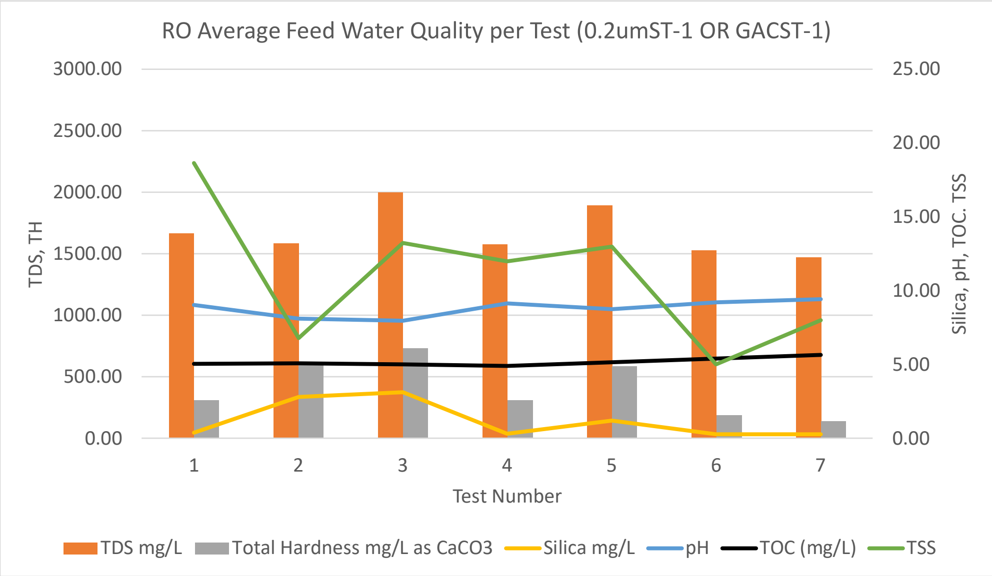
29% reduction in total hardness likely because of the elevated pH, whereas Test No.7 only had a 5% reduction in total hardness indicating the concentrate recycle to the EC had an appreciable total hardness reduction in Test No.6 TDS but not in Test No.7 and the only thing different between the two tests was the concentrate recycle.

TSS: All of the tests had significant TSS increases between 5800% to 13,300% increase because of concentrate recycle, with the exception of Test No.2 that did not have a concentrate recycle and no change in TSS because it was the same sample as the one collected at RWST-1. Test No.3 only had a 30% increase in TSS.

Silica: Test No.6 had the highest silica reduction at 82% followed by Test No.1 at 66%, Test No.7 at 41% and Test No.5 at 21%. This is counter-intuitive of the solubility relationship with silica and pH according to Section 4.2.1.

pH: The pH remained unchanged at the ambient pH test runs No.2 and No.3. A 15% to 26% increase in pH at the RWST-2 tap was noticed for all the remaining tests where caustic was added, with the concentrate return to the EC test having the higher increase in pH as expected.

TOC: Regardless of the test run, the TOC concentrations between RWST-1 and RWST-2 remained unchanged.



Test #	Testing Constraints	# of Samples Analyzed
1	10.3 pH/CON to EC/1 min	8
2	AMB pH/ No CON Recovery/1 min	4
3	AMB pH/CON to EC/1 min	4
4	10.3 pH/CON to Clarifier/1 min	2
5	10.3 pH/No CON to EC/1 min	2
6	10.3 pH/CON to EC/3min	1
7	10.3 pH/No CON Recovery/3min	2

Figure 28.—RO average feed water quality per test (0.2umST-1 OR GACST-1).

Results presented in this section address the impact EC and particle separation had on the RO performance, since this grab sample was representative of what was feeding the RO system.

RWST-2 to RO Feed (0.2umST-1 or GACST-1) Trends

TDS: Tests No.1, No.3 and No.6 had a 13% to 22% reduction in TDS likely because it efficiently removed the hardness portion of the TDS; however, Test No.2 and No.7 only saw a marginal reduction in TDS. Tests No.4 and No.5 were the only tests to see a 4 to 24% increase in TDS, respectively, likely because caustic was added, and concentrate was returned to the clarifier in Test No.4. By the following day when Test No.5 was initiated the higher TDS volume of the clarifier was pumped through the system and was captured in the grab samples that day.

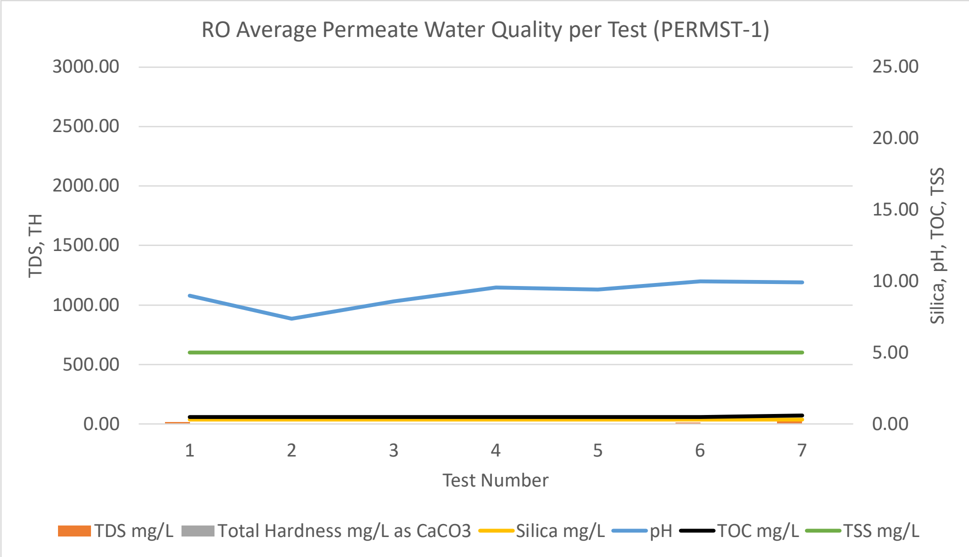
Total Hardness: All of the tests had a significant additional total hardness reduction ranging from 23% to 80% with Tests No.2 and No.3 having the least additional total hardness removal at 23 and 24%, respectively, because they were operated at a 1 min HRT at ambient pH. The additional total hardness reduction from the RWST-2 sample tap is likely because the precipitated pinfloc was removed through the cartridge filters. The total hardness concentrations from this sample tap, which was the water feeding the RO membranes, ranged between 140-mg/L in Test No.7 to 735-mg/L in Test No.3, with no pretreatment other than EC + cartridge filtration.

TSS: Tests No.1, No.4, No.5 and No.6 had a 95 to 99% TSS reduction because of the pinfloc removal efficacy of the cartridge filters and the efficiency of the larger floc diameter either by longer HRTs or caustic addition. However, Test No.3 had a 104% increase in TSS indicating the floc size was smaller than the 0.2-micron rating of the filter and the ambient pH at 1-min HRT. This coupled with concentrate recycle adversely affected the performance by not allowing a large enough floc particle to either settle effectively or be filtered through the prefiltration systems.

Silica: Regardless of pH or HRT, the EC unit removed between 78 to 97% of silica, even at pH ranges between 9.7 and 10.3, where amorphous silica has a solubility concentration greater than 800-mg/L at 22ºC. See Section 4.2.1 for a more detailed explanation of this phenomenon. See Section 4.2.1 for a more detailed explanation of this phenomenon.

pH: In all tests, except Tests No.2 and No.3 that were ran at ambient pH, the pH increased to between 10.3 and 11 as expected due to the addition of caustic.

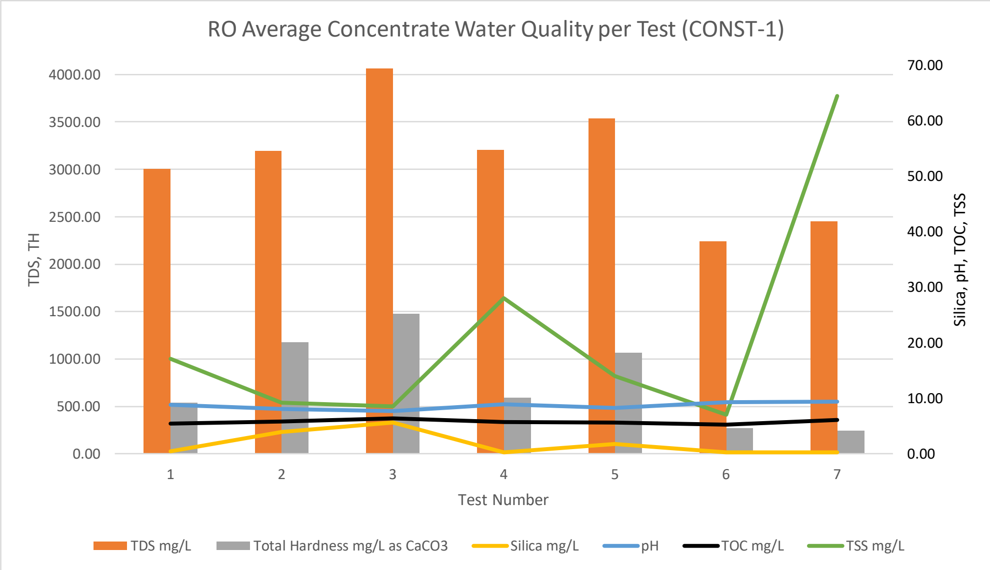
TOC: All test runs had a 10 to 24% reduction in TOC levels, likely because of the tighter filtration prior to the RO feed was able to remove the e-floc that had been generated.



Test #	Testing Constraints	# of Samples Analyzed
1	10.3 pH/CON to EC/1 min	8
2	AMB pH/ No CON Recovery/1 min	4
3	AMB pH/CON to EC/1 min	4
4	10.3 pH/CON to Clarifier/1 min	2
5	10.3 pH/No CON to EC/1 min	2
6	10.3 pH/CON to EC/3min	1
7	10.3 pH/No CON Recovery/3min	2

Figure 29.—RO average permeate water quality per test (PERMST-1).

The water quality leaving the RO as permeate was essentially distilled water quality as all the previously trended analytes were non-detect including the average TDS concentrations ranged between 6.2-mg/L and 18.5-mg/L. The membrane never fouled after a brief permeate rinse after each run. Later in the FMA, there the recovery was increased to 97% with occasional concentrate purge to the VAED system. In lieu of a detailed water quality analysis, the inline conductivity meter was used to track treatment efficacy. Conductivity in the raw water was consistently removed over 99.2% in the permeate, while specific flux was maintained after a 10 to 15-gallon permeate rinse.



Test #	Testing Constraints	# of Samples Analyzed
1	10.3 pH/CON to EC/1 min	8
2	AMB pH/ No CON Recovery/1 min	4
3	AMB pH/CON to EC/1 min	4
4	10.3 pH/CON to Clarifier/1 min	2
5	10.3 pH/No CON to EC/1 min	2
6	10.3 pH/CON to EC/3min	1
7	10.3 pH/No CON Recovery/3min	2

Figure 30.—RO average concentrate water quality per test (CONST-1).

The water quality leaving the RO as concentrate had all the dissolved ions at the correct concentration factor that was larger than the molecular weight cut-off of the membranes, if they were not already removed in the pretreatment.

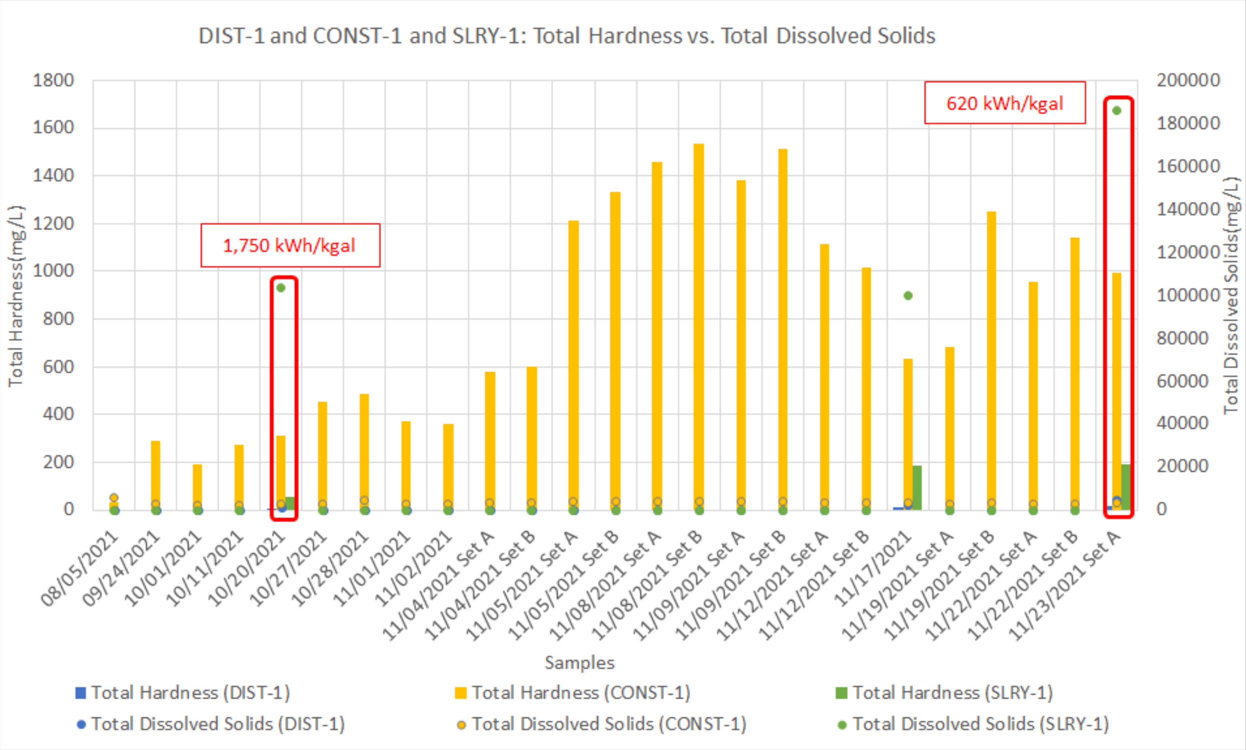


Figure 31.—VAED lab analysis for the three grab sample sets vs the concentrate concentrations.

The VAED system “electro-distilled” the concentrate collected in the CST but required salt doping and salt slurry re-seeding for it to operate within its design constraints as described in Section 2.3.3. Since the VAED system operated as a batch process, only three grab samples were sent to the lab for analysis on October 20, 2021, November 17, 2021 and 11/23/2021. The VAED system removed 98% to 99% of the TDS for three lab analysis dates where the raw TDS was 186,060 mg/L, 100,013 mg/L and 103,531 mg/L at an energy intensity between 620 to 1,745 kWh/kgal to create up to 5 gph of high-quality distillate. Refer to Figure 31.—VAED lab analysis for the three grab sample sets vs the concentrate concentrations that reveals the lower energy intensity occurring from the highest TDS brine, confirming the efficiency gains from ohmic heating in the flash chamber.

As brine evaporated and was condensed into distillate, the brine level decreased in the flash chamber until it eventually dropped below the bottom of the electrode and the circuit turned off. As the conductivity of the brine increased, while the pure distillate was evaporated, the ohmic heating impact of the brine circuit caused the boil aggressiveness to significantly increase. The energy intensity numbers in the above graph were for the obsolete cooling mechanism that could not keep up with the rate of steam generation as the conductivity increased from evaporation. The heating element rheostat had to be turned down, so the steam generation rate matched the rate at which the chiller was able to condense the steam; otherwise, the vacuum would be lost, and the boiling point would increase. The max temperature manually set in the flash chamber was 150ºF to prevent the melting of the PVC flash chamber.

To fully capture the efficiency gains as conductivity increases, additional cooling with a more efficient heat exchanger was incorporated in the VAED system the same time the FMA mode was initiated for the main process train. Once the additional cooling was implemented, steady state operation at 5-gph was achieved as the brine concentrations reached close to saturation. Details of the upgrades will not be disclosed in this research because of the proprietary nature of the patent.

The following section attempts to discuss the impact each major unit process had on the treatment efficacy both during the pilot phase and Failure Mode Analysis.

4.2.1 Chemical Pretreatment

- Increasing the pH with sodium hydroxide from 8.6 to between 9.7 and 10.3 resulted in total hardness (TH) removal efficacy prior to the RO between 59% to 80% in the presence of antiscalants; however, the data obtained did not permit a more detailed analysis of what fraction of hardness removal was because of pH change and what fraction was because of the EC treatment because:
    - A test run was not performed with the EC blades de-energized, which would have provided insight as to the fraction removed by increasing pH through caustic versus EC.
    - An acid preservative was added to sample bottles to facilitate ICP measurement of metal concentrations. As a result, the hardness measurements indicated dissolved and some redissolved hardness.
- Without EC, at a pH range between 9.7 and 10.3, the solubility of amorphous silica should have resulted in no reduction in the average silica concentrations of 17.7-mg/L in the blowdown. Referring to Figure 32.—Solubility graph of amorphous silica vs pH the solubility of amorphous silica in 22ºC (71.6ºF) water at a pH of 10, which is representative of the pilot trailer operation, increases to approximately 800-mg/L. The EC step resulted in silica removal between 83% and 98% at EC residence times between 1 minute and 3 minutes, regardless of the pH, even at the higher pH's previously mentioned. This is likely due to the formation of iron and aluminum silicate floc from the EC.

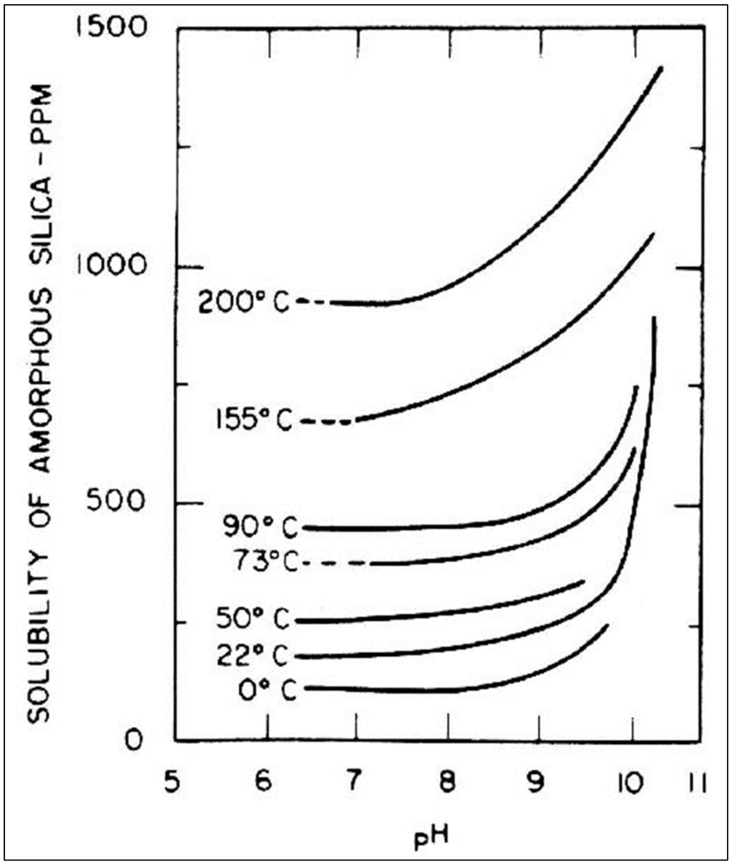


Figure 32.—Solubility graph of amorphous silica vs pH.

4.2.2 Electrocoagulation

- All EC removal efficacy numbers and trends in this section compare the raw water to the 0.2-micron filtrate when the EC was energized with and without pH adjustment.

- EC test runs that operated at ambient pH of 8.6 only experienced TH removal between 23% and a 2% increase depending on if concentrate was recycled or not.

- Typically, no more than 170-mg/L to 256-mg/L as CaCO3 TH is recommended in the RO feedwater without proper pre-treatment with antiscalants or pH adjustment to prevent irreversible membrane fouling. An average TH concentration feeding the RO for all test runs was 428 mg/L as CaCO3 with a maximum TH of 735 mg/L as CaCO3 in Test No.3, which was ambient pH, concentrate recycled to EC at a 1-minute hydraulic residence time. No fouling or process degradation was experienced.

- The electrocoagulated floc was significantly diminished after approximately 4 to 6 hours (360 to 540 gallons) because of scaling of the electrodes and required frequent CIPs with 34% hydrochloric acid (HCl).

- Because of the high hardness concentrations in the blowdown, polarity reversal frequency of the EC unit needed to be reduced from once per minute to four times per minute to extend runtime between CIPs and prevent scaling. Runtimes were increased by approximately 30% after increasing polarity reversal frequency.

- The CIP HCl was recycled between 8 to 11 times until the pH increased > 3, where it was neutralized and disposed. Newer EC units have capabilities of processing the spent CIP solution.

- In all tests, EC depressed the average ORP from 194 mV to -112 mV indicating a reducing condition and the absence of oxidizers that would harm the membrane.

- Regardless of pH or HRT, the EC unit removed between 83 to 98% of silica prior to the RO, even at pH ranges between 9.7 and 10.3, where amorphous silica has a solubility concentration greater than 800-mg/L at 22ºC.

- In all tests, EC depressed the average ORP from 194 mV to -112 mV indicating a reducing condition and the absence of oxidizers that would harm the membrane. The negative ORP, and the fact that EC floods the water with negatively charged electrons, may also play a significant role in the antiscalent effect of the EC as explained later in the FMA Section.

- An average current of 5 amps was applied during the field testing which resulted in a large sweep floc that was instantaneously formed in the reaction chamber. During the FMA, the electrodes were reconfigured to split the chamber into two equal sections as opposed to only one chamber. The amperage increased to 20 amps and the floc formed was much denser. The higher amperage configuration was operated for over 15 hours (and still continuing) without requiring a CIP. Currently, 18-months later, the we only performed one CIP, not because the RO was underperforming but because we wanted to show a tour attendee what the dense e-floc looked like with clean blades.

- Total coliform samples were collected throughout all locations for every grab sample event. Only one sampling event came back with a 4-most probable number (mpn)/100 mL value for the raw blowdown on 11/1/2021, while all the other locations and dates were below detection limits of 0.5-mpn/100mL. The corresponding ECST-1 sample tap, which was after EC but before the 20-micron filter, was non-detectable. This one data point is not enough to confirm the electrophoresis disinfection effect; however, it is safe to assume that later during the FMA, there was considerable biological activity because of the significant algae growth that can be seen in Figure 6.—Picture of blowdown

- capture pool during failure mode analysis (FMA) with significant algae growth The RO system was not fouled after being electrocoagulated with a 1-min HRT, likely proving the EC can prevent biological fouling.

4.2.3 Clarification

- The custom fabricated underflow clarifier did allow some floc carry over at a 1-min HRT (1.5 gpm) causing noticeable build-up in the 20-micron cartridge filters but not on the 0.2-micron filters. There was no e-floc carryover with the 3-min HRT (0.5 gpm). The adjustable sleeve was not moved up to allow a greater cross-sectional area for the solids laden water to flow as the floc level increased in the annular space between the shroud and the cube. This likely led to excessive fluid velocities at the exit of the shroud that caused the bed to rise and re-released the settled particles. Being more vigilant about adjusting the shroud level may have reduced the pinfloc rise and prolonged cartridge filter runtime.

4.2.4 Microfiltration + Reverse Osmosis

- The MF system consisted of a 20-micron cartridge filter followed by a 0.2-micron cartridge filter during the pilot tests. The 20-micron filter was replaced with a 5-micron cartridge filter during the FMA testing. Regardless of the prefilter configuration, a 0.2-micron filtrate was fed to the RO boost pump.

- The GAC system did not provide TOC removal and was not a benefit to the process train. In fact, an increase in total suspended solids (TSS) was noticed in the GAC runs likely because of sloughing, so this configuration of testing was immediately stopped so as not to potentially foul the membrane.

- The 2-stage DOW FilmTec^{TM} seawater RO membrane system initially used in the pilot was replaced with a 3-stage system during the first month of operation after a test run significantly exceeded the maximum flux rate, causing the membranes to delaminate and fail. Adding the 3rd stage and throttling the 1st stage permeate valve maintained proper crossflow velocity and higher recoveries. However, using multiple arrays in the 1st two stages and using brackish water RO versus seawater RO would likely increase the recovery and decrease the energy intensity.

- The RO consistently removed 99.3% of the TDS and 99.7% of the total hardness from the electro-coagulated MF filtrate for all 7 tests. The average electro-coagulated MF filtrate feeding the RO had 1,686-mg/L TDS and 429-mg/L total hardness as $CaCO_3$.

- The manufacturer recommended specific flux was 0.030 gfd/psi while all test runs experience specific fluxes between a 10th percentile of 0.040 to a 90th percentile of 0.050 gfd/psi. After each test run was completed the specific flux and TDS removal efficiency was recovers with a 10 to 15 gallon permeate flush.

4.2.5 Vacuum Assisted Electro-Distillation Concentrate Treatment

- The VAED system removed 98% to 99% of the TDS for three lab analysis dates where the raw TDS was 186,060 mg/L, 100,013 mg/L and 103,531 mg/L at an energy intensity between 620 to 1,745 kWh/kgal to create up to 5 gph of high-quality distillate.

- Higher conductivity brine feed to VAED caused a more aggressive boil in the flash chamber (because of ohmic heating) than anticipated. This resulted in the 5,200 BTU/hr chiller not being able to keep up with condensing the steam at the same rate it was generated, which results in a loss of vacuum and increased boiling point. To prevent melting the PVC flash chamber, the power supplied to the heating element across the brine circuit was reduced, which decreased the distillation efficacy. Increased chilling BTU/hr rate will help increase the distillation efficiency.

- As the brine in the flash chamber reaches a certain conductivity, motor loads can be placed on the brine circuit and the resistive heating element can be removed. This will further increase the distillation efficacy by introducing multiple effects in addition to the depressed boiling point from the vacuum.

- A 12,000 BTU air conditioner and used radiator was integrated into the cooling loop of the VAED recirc water to help enhance the cooling effect on the distillation hood. This allowed the unit to attain steady state operation by increasing the condensation rate, so it matched the steam generation rate. Under this configuration the VAED attained 5-gph distillation rate between 100°F and 120°F at a 21-inch Hg to 23-inch Hg vacuum on greater than 180,000 mg/L TDS salt seeded concentrate.

- The standard 1.5-kW, zinc coated resistive heating element showed no signs of corrosion or scaling after sitting submerged in 100,000 to greater than 200,000 mg/L brine for over 6 months, while being periodically operated 16 to 24 hours a month.

4.2.6 Failure Mode Analysis

After field testing was complete in November 2021, the research team began the FMA by recirculating the concentrate to the EC in a proprietary manner with no chemical pretreatment to the RO, only EC + CLAR + MF. After producing between 480 to 500-gallons of permeate (over weeks of start stop operation and no pickling of the membranes), the feed pressure rate of gain started to increase from 600-psi and the permeate flow decreased. The operating pressure of the RO was 800-psi, with a pressure relief set at 1,000-psi. This initiated a 10 to 15 gallon concentrate purge to the VAED system. In the initial phases, this was followed by a 10 to

15 gallon permeate flush. However, later in the FMA the permeate flush was not initiated, and the specific flux and TDS removal efficiency consistently recovered to values detailed in table E-1.

Below are some more trends noticed during the FMA:

1. Throughout the FMA, the electrocoagulated, 0.2-micron filtrate was left in the RO elements without adding sodium bisulfite preservative for one to three months with no apparent and irreversible fouling after a purge procedure, detailed in the previous paragraph, was initiated.

2. There was not a laboratory water quality analysis during the FMA, only surrogate measurements from inline analyzers as previously mentioned.

3. The TH concentration feeding the RO was estimated to be at least 4 times that of the pilot phase described in table ES-1 based on the conductivity increase from the concentrate recycle as read from the inline conductivity meters, with no apparent fouling.

4. An EC CIP was only initiated one time during the FMA prior to a pilot trailer tour to show attendees the abilities of EC in generating a dense, heavy floc.

5. In all tests, EC kept the ORP feeding the RO to below –165-mV indicating a reducing condition and the absence of oxidizers that would harm the membrane.

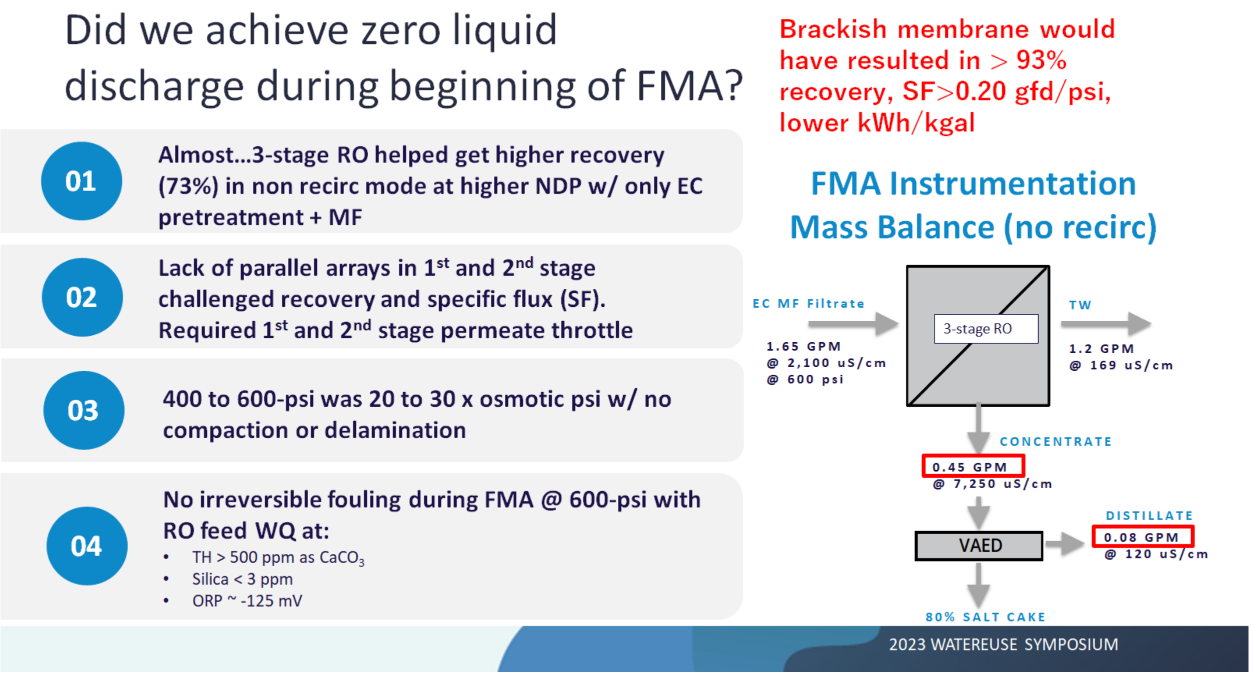
6. As of January 2024, the system continues to operate in FMA with no pretreatment other than EC at a 1-minute HRT with fouled electrode surfaces. No membrane cleanings occurred, nor was disinfectant or caustic added to pilot system since the pilot began in August 2021. There was no apparent biological growth in the unit processes downstream of the EC such as the clarifier, clear cartridge filter housings, RO elements, concentrate storage tank and permeate storage tank (refer to Appendix A-2 for an explanation from previous research by University of South Florida and Powell Water). Post treatment of the permeate was not performed with respect to pH adjustment or disinfection for all phases of testing,

7. The VAED system continued to remove 98 to 99% of the TDS from the seeded concentrate as indicated by the conductivity tester in the trailer. The rate of distillation consistently reached steady state at 120ºF to 130ºF at a up to 18-gph; however, as the conductivity increased to approximately 65,300-uS/cm, the temperature needed to be decreased because the chiller could not keep up with the flash rate which caused salt laden foam to migrate into the heat exchanger and ultimately into the distillation trough. Salt laden foam in the distillation trough would contaminate the sub-100-uS/cm distillate and require a permeate flush of the distilled water system.

8. The standard 1.5-kW, zinc coated resistive heating element showed no signs of corrosion or scaling after sitting submerged in 100,000 to greater than 200,000 mg/L brine for over 30 months, while being periodically operated 16 to 24 hours a month.

Below are several figures taken from Garver's Presentation at the WateReuse 2023 Symposium in Atlanta, GA on March 6, 2023, titled "Lessons Learned from a BOR Funded Desalination Research Project for Cooling Tower Blowdown" that explains the mass balance of the FMA as it relates to how the treatment train was able to get to ZLD.

The explanation of why the RO membranes were able to consistently achieve 35% higher specific flux than recommended by DOW may be explained in Dr. Kamyar Sardari's post doc studies at the University of Arkansas (citation No.15 in the reference section) focused on researching bench top, atmospheric EC technology in the pretreatment role in UF and forward osmosis (FO) membranes on challenged waters. According Dr. Sardari's membrane autopsy in Figure 34, EC clearly had a positive "anti-fouling" impact on the FO membrane surface as opposed to the control FO membrane that did not implement EC as membrane pretreatment. The longer the EC HRT, the cleaner the FO membrane surface.



Did we achieve zero liquid discharge during beginning of FMA?

Brackish membrane would have resulted in > 93% recovery, SF>0.20 gfd/psi, lower kWh/kgal

01 Almost…3-stage RO helped get higher recovery (73%) in non recirc mode at higher NDP w/ only EC pretreatment + MF

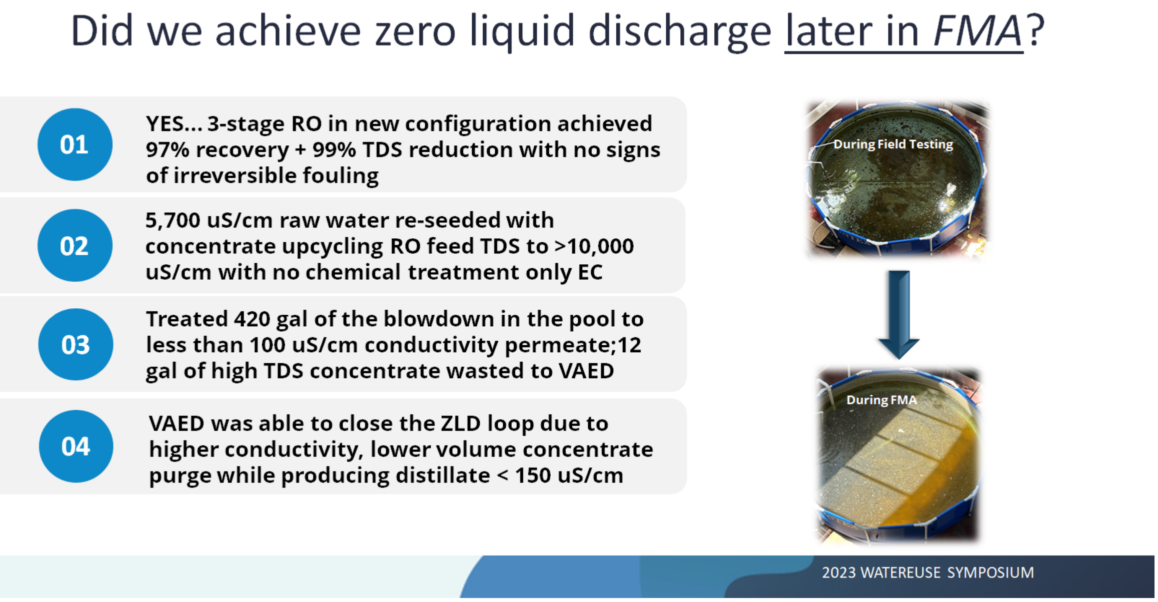
02 Lack of parallel arrays in 1st and 2nd stage challenged recovery and specific flux (SF). Required 1st and 2nd stage permeate throttle

03 400 to 600-psi was 20 to 30 x osmotic psi w/ no compaction or delamination

04 No irreversible fouling during FMA @ 600-psi with RO feed WQ at:
- TH > 500 ppm as $CaCO_3$
- Silica < 3 ppm
- ORP ~ -125 mV

FMA Instrumentation Mass Balance (no recirc)

2023 WATEREUSE SYMPOSIUM



Did we achieve zero liquid discharge later in FMA?

01 YES… 3-stage RO in new configuration achieved 97% recovery + 99% TDS reduction with no signs of irreversible fouling

02 5,700 uS/cm raw water re-seeded with concentrate upcycling RO feed TDS to >10,000 uS/cm with no chemical treatment only EC

03 Treated 420 gal of the blowdown in the pool to less than 100 uS/cm conductivity permeate;12 gal of high TDS concentrate wasted to VAED

04 VAED was able to close the ZLD loop due to higher conductivity, lower volume concentrate purge while producing distillate < 150 uS/cm

2023 WATEREUSE SYMPOSIUM

Figure 33.—FMA mass balance explanation as it relates to zero liquid discharge (ZLD) metric.

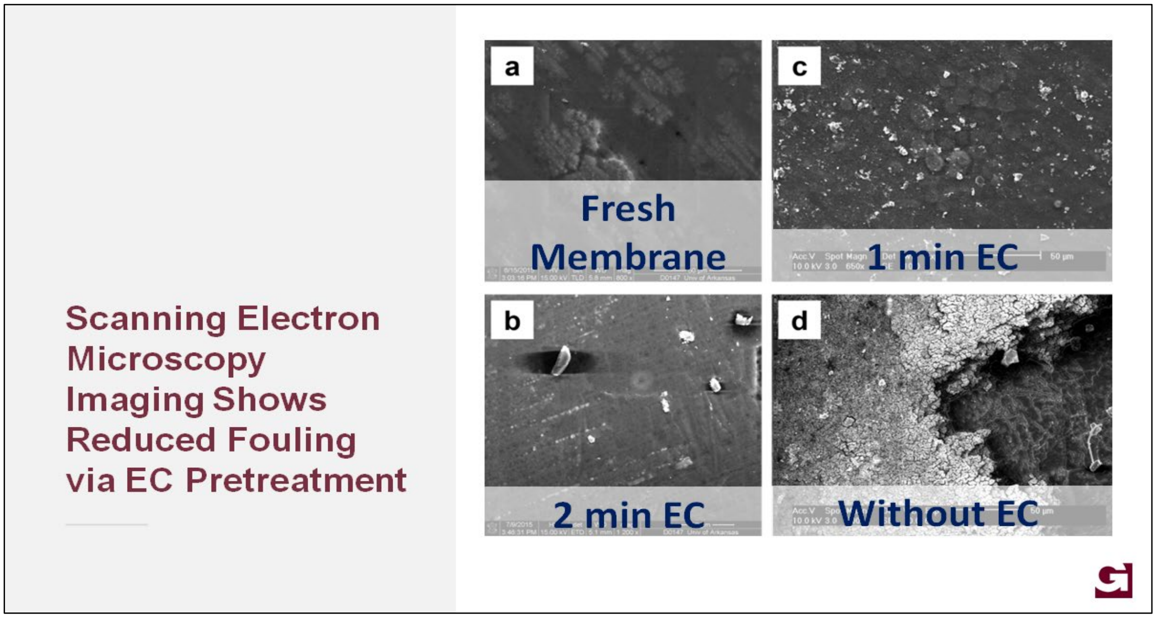

Scanning Electron Microscopy Imaging Shows Reduced Fouling via EC Pretreatment

Figure 34.—Scanning electron microscopy of FO membrane for aluminum EC + FO for Treating Produced Waters (2018), Desalination. (428) pp. 172–181.

In another of Dr. Sardari’s research projects on produced water samples from a natural gas extraction facility in Marcellus shale, his research indicates EC pretreatment reduces the rate of flux decline in membrane distillation indicating fouling suppression, where experiments continued for 400+ hours with no reduction in flux as can be seen in figure 35.

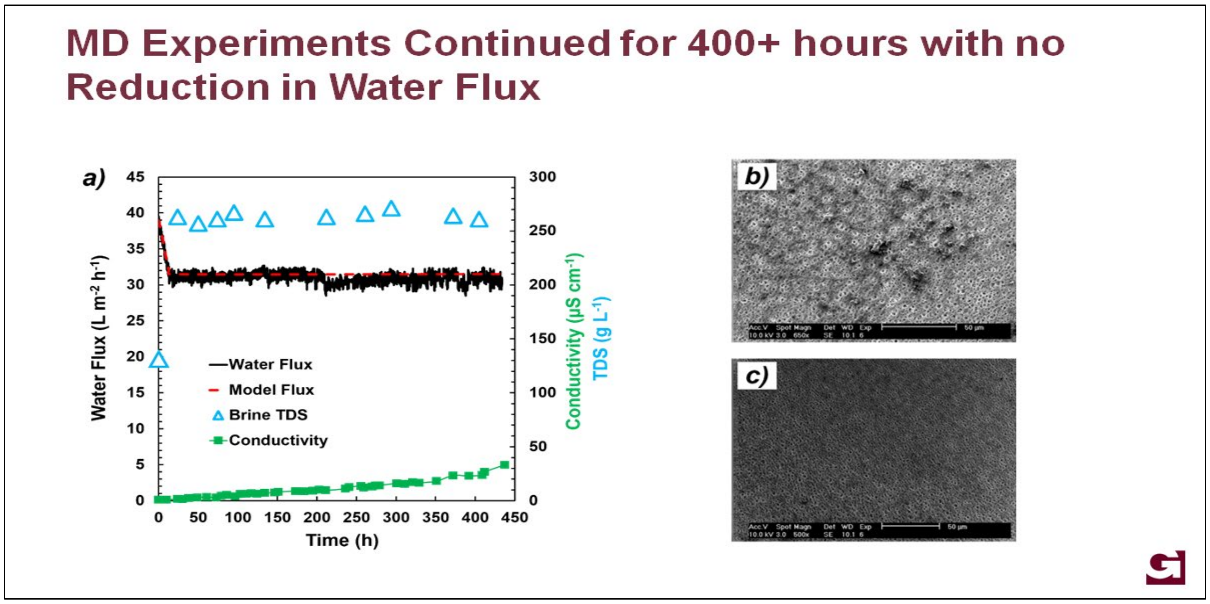
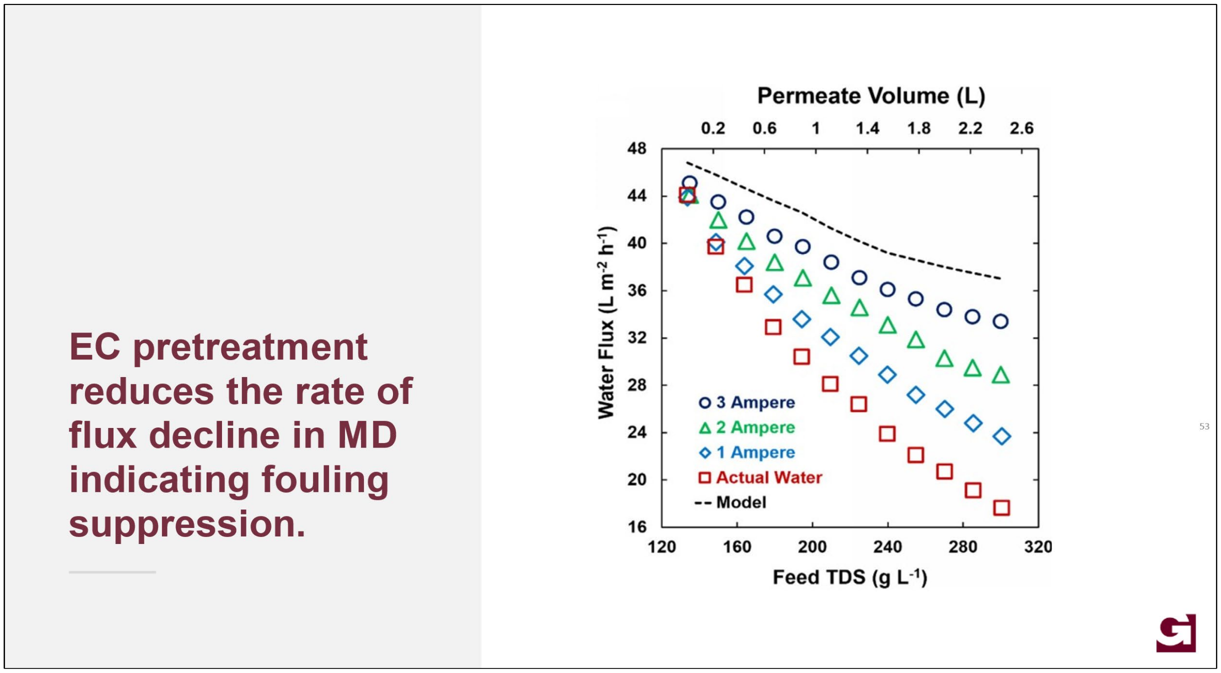

MD Experiments Continued for 400+ hours with no Reduction in Water Flux

Figure 35.—Membrane flux as a function of increasing TDS and operating time on produced waters (2018), Desalination. (428) pp. 172–181.

Dr. Sardari's research corroborates the research in this project that the rate of flux decline reduces as higher amperage is passed through the liquid as can be seen in figure 36.



EC pretreatment reduces the rate of flux decline in MD indicating fouling suppression.

Figure 36.—Membrane flux as a function of increasing TDS and amperage on produced waters (2018), Desalination. (428) pp. 172–181.

5.0 Conclusions

5.1 Conclusions

Proof of concept was achieved. There is a benefit of utilizing atmospheric EC ahead of membranes; however, there are opportunities for improvement. The pilot train was operated in two configurations on the saline cooling tower blowdown that ranged between 1,800-mg/L TDS during the pilot to greater than 15,600-mg/L TDS during the FMA. The system consistently and effectively produced high-quality permeate less than 20-mg/L TDS in all phases of operation without fouling. The system was only operated in a start stop operation with periods of days to 3 months between operation without a RO membrane CIP or pickling routine performed.

In the optimal foulant removal (OFR) mode, the EC/Clarification/MF train removed target RO foulants such as total hardness and silica. This mode incorporated a 3-minute HRT and/or 9.7 to 10.3 pH and achieved up to 80% total hardness removal and 98% silica removal at a 0.04 to 0.05-gfd/psi specific flux. 50% to 54% recovery was experienced in the OFR mode with no fouling, compaction, or delamination. An average pressure of 346-psi was applied in these tests. In the OFR mode, it was assumed the RO concentrate was suitable for sodium hypochlorite generation, after being distilled in the VAED system, since there was insignificant hardness and silica in the VAED salt slurry because it was removed earlier in the process with the EC and particle separation step. A higher brine quality (without hardness and silica) can potentially be used as a feedstock for the chlor-alkali market to generate sodium hypochlorite onsite. This sodium hypochlorite may be able to be used back in the cooling tower for disinfection purposes. This would help promote a circular economy; however, further research is needed to determine if the quantity and quality can consistently meet the needs of the onsite generation system.

In the moderate foulant removal (MFR) mode, conducted during the FMA, the EC/Clarification/MF system was operated at a stressed condition with only a 1-min HRT and ambient pH of 8.3, while the RO was operated at an average of 600-psi, a specific flux value at 0.046 gfd/psi and a 73% recovery with no fouling, compaction, or delamination. The 600-psi setting in the FMA was roughly thirty times the osmotic pressure requirement and there was no signs of compaction or delamination. Although this mode likely had high hardness concentrations in the concentrate due to the concentrate recirculation, based on previous test runs, making it unsuitable to be used as a feed stock for onsite sodium hypochlorite generation, it did not appear to foul the RO system. In fact, one of the highest recoveries were noted in the non-optimal FMA mode.

Zero liquid discharge was not achieved during the pilot testing, because of the lower recovery of the membrane system configuration resulted in a lower conductivity, increased concentrate flow of 0.45 gpm, and the VAED system was only capable of processing 0.08-gpm. However, the highest recovery the pilot system attained was during the FMA as described in Section 4.2.6 – Failure Mode Analysis where the RO recovery was increased from approximately 54% to more than 97%. During the FMA, the pilot was operated in a concentrate recirculation mode with intermittent operation without membrane CIPs or sodium bisulfite preservative procedures implemented during prolonged downtimes of up to 4 months. The reduced amount of high saline concentrate generated during this mode of operation (i.e.; MFR mode) allowed the VAED system to keep up with the concentrate generation rate; therefore, closing the ZLD loop.

With a brackish water membrane configuration and multiple parallel arrays in the 1st two stages, the research team is confident the RO can achieve between 92% to 95% recovery at a specific flux greater than 0.2 gfd/psi. This recovery results in 0.08-gpm (5-gph) of high TDS brine to the VAED, which had a 5-gph flow rate when the brine concentration exceeded 180,000 mg/L, so true ZLD can be achieved.

The custom clarifier and MF system used to settle and filter the EC floc met the treatment goals. Silica removal efficiencies consistently exceeded 98% regardless of treatment configuration resulting in RO feed concentrations between 0.1 and 3 mg/L from an average raw water concentration of 17.7 mg/L. The GAC system did not provide TOC removal and was not a benefit to the process train. In fact, an increase in TSS was noticed in the GAC runs likely because of sloughing. Although an RO autopsy was not performed, it can be assumed that biological fouling was also prevented because of the EC and its ability to achieve a 3 to 5 log bacteria and virus kill per Dr. Mya Breitbart, University of South Florida. See Appendix A-2. This was proven by the RO system not fouling after leaving electrocoagulated, 0.2-micron filtrate in the modules for over 2 months without preservatives or flushing.

The ohmic heating impact on efficiency on the VAED system was confirmed during testing. The rate of heat generation will increase as the electrical conductivity increases because of the increasing concentration as the seawater evaporate.[14] The existing capacity of the VAED system would have been high enough to close to achieving ZLD, if the RO system was able to achieve a higher percent recovery. As the conductivity increases, the resistive heating element can be removed, and the motor loads can be placed on the brine circuit to increase efficacy by adding multiple effects.

With automation, enhanced RO recovery and increased cooling efficiency of the VAED system, the EC/Clarification/MF/RO with VAED treatment train has potential of achieving zero liquid discharge (ZLD) treatment for the blowdown from the 600-ton cooling tower at RRCC at a 1.5 gpm flow rate; however, because of some limitations, minimum liquid discharge was achieved. Automation of the treatment train would also allow for longer, unstaffed runtimes so the system can be tested for more than five to six hours at a time. More research into the removal mechanisms, optimization of full-scale design constraints, capital costs, and operational costs would be beneficial to the advancement of this technology as a solution to manage the salt loading from cooling towers, brackish water desalination, and potentially seawater desalination applications. Additionally, in future phases, more research should be focused on the electro-anti-scalant phenomenon, and the fate of the residuals generated from the process.

5.2 Challenges

The $200,000 grant served its purpose of vetting the viability of this technology as an emerging opportunity to enhance RO pretreatment and manage the concentrate. However, commercialization of this technology requires substantially more analytes tested (and tested more frequently), optimized RO and VAED configurations and automation of the entire treatment train. The lab analysis, engineering, and research required to design and build these improvements, confirm removal efficacy, confirm design constraints, operation costs and capital costs could easily exceed $2,000,000.

It is important to note that after testing was completed, the research team realized an additional test could have been conducted for any one of the above tests where caustic was used to increase the pH. This additional test would have the same constraints in place; however, the pH adjusted water would have passed through the EC reaction chamber with the electrodes de-energized. This type of test would act as a “control” and establish a baseline removal efficiency of target foulants from the contribution of caustic addition. This test can be compared to the contribution EC had to the increased foulant (i.e., silica, and hardness) removal efficacy of the pretreatment system.

The hardness removal efficiency trends in this pilot aligns with past experience the research team has on similar water quality where EC alone tends to remove 40% to 60% of the total hardness at pH between 6.5 and 8.7. Increasing the pH above 9.4 typically results in 90 to 97% removal efficiency.

5.3 Hypothesis Generated from Research

Our team developed several hypotheses from this research as follows:

- There is an “antiscalant-like” phenomenon of irreversible fouling because of the EC unit is as follows. The EC floods the water with negatively charged electrons which causes a negative ORP and a slight negative charge on the water. Since the polyamide RO membranes are also negatively charged, a repulsion effect of common divalent, positively charged scaling components was noticed.

- The direct current electric field induced on the water in the EC reaction chamber had an electrophoresis effect on biological activity as corroborated by other research identified in Appendix A-7. Only a surrogate comparison could be achieved in this research by the fact that the membranes never irreversibly fouled under biological growth in the blowdown water. Furthermore, nothing appeared to grow in the membrane modules during excessive down time without preservatives or pickling agents since the specific flux, percent recovery and percent salt removal was consistently recovered.

- The VAED system combines vacuum distillation with ohmic heating to flash the saline brine between 120°F and 135°F in 8-gallon batches at 5-gph. As the water evaporated, the conductivity increased as did the aggressiveness of the boil as the amps increased, while the rheostat that controlled the amperage across the brine circuit was held constant. The contribution of ohmic heating with vacuum distillation was apparent as the boil quickly subsided when the brine circuit was de-energized, while the temperature and vacuum level in the flash chamber remained constant. The rate of condensation must match the rate of steam generation or the vacuum will decrease and the boiling point will increase.

5.4 Recommended Next Steps

5.4.1 Fully Automate the Treatment Train

With automation, enhanced RO recovery and increased cooling efficiency of the VAED system, the EC/Clarification/MF/RO with VAED treatment train will allow an increased ease of operation of the ZLD treatment train and increased flow capacities. More research into the removal mechanisms, optimization of full-scale design constraints, capital costs, and operational costs would be beneficial to the advancement of this technology as a solution to manage the salt loading from cooling towers, brackish water desalination, and potentially seawater desalination applications. Additionally, in future phases, more research should be focused on the electro-anti-scalant phenomenon.

5.4.2 Analyze the Solids from the Treatment Process for Beneficial Reuse as Feedstock for Onsite Sodium Hypochlorite Generation Potential

There was no budget remaining to evaluate the quantity and quality of solids generated from the pretreatment process and the salt slurry residual from the VAED system. Analyzing the residuals from a treatment process, especially in desalination, is critical to developing a techno-economic analysis to support the design. If the salt slurry leaving the VAED system had high purity NaCl with no foulants such as hardness or silica, it may be used as a feedstock for onsite sodium hypochlorite generation that can be used in the cooling tower per previous discussions.

5.4.3 Compare EC Versus Electro Dialysis Reversal (EDR) As Membrane Pretreatment on Challenged Waters

There is a big push in the desalination industry for electrified desalination treatment. EDR is another type of treatment that falls in the electrified category because it uses direct current polarity reversal to regenerate a charged anion and cation membrane once the membrane surface is passivated. The EDR technology has made some recent improvements that allow for higher throughputs, higher efficiencies, and safer operation than some of the previous versions. It would be beneficial to see how EDR compares to EC as a membrane pretreatment with identical membrane treatment trains located downstream of each system.

5.4.4 Prove Electrocoagulation Antiscalant Effect Hypothesis on RO Membrane Surface

If EC has a true "electric" antiscalent effect on a polyamide RO membrane, it could be significant to the desalination industry. This claim requires further research to develop feasibility and a techno-economic analysis. A post research RO membrane autopsy is recommended on all 3 stages to help reveal the level of fouling, similar to what was performed in Dr. Sardari's membrane autopsy in figure 34.

5.4.5 Determine Exact Contribution of Electrical Current Passed Through the Water in Destroying Biological Activity

If EC does have a virus and bacteria kill, it could have a profound impact on brackish water and seawater desalination technology. The electrophortesis that occurs because of EC may be effective at preventing biofouling of the membrane surfaces as previously described. Bioassays should be conducted to quantify the log inactivation for common water-born bacteria and viruses, along with identifying the mechanism of destruction. E. coli is a commonly used bacteria surrogate and bacteriophage MS2 is a common virus surrogate in the water treatment processes due to its easy detection methods.

5.4.6 Determine Exact Contribution of Caustic Addition to Hardness Removal

If using the existing pilot trailer to determine the contribution of caustic addition to the hardness and silica removal, the exact same caustic dose rate should be used with the EC reaction chamber de-energized. This will be a true comparison and allow the researcher to determine the impact pH adjustment with caustic has on the treatment efficacy. It is recommended that this test be performed as the last, since the anti-fouling contribution of EC may not be realized for this test and the RO may foul if considerable hardness and silica concentrations enter the membranes. For a bench top analysis to prove the same concept, please refer to Appendix A-7

5.4.7 Further Research the VAED Technology for More Aggressive Brine Solutions

The VAED system distills better as conductivity increases due to ohmic heating. Spent brine from ion exchange resins and water sources such as the Paradox Valley and Salton Sea brine sources may be a good source to research the technology. The BOR already did a paper evaluation of Brine Bulb Technology (aka VAED) back in 2016 in a technology assessment titled “Paradox Valley Unit Brine Crystallization Technology Assessment”. Now that the VAED technology proved its efficacy in treating high TDS water in this research, it may be a good candidate for research in these challenged waters.

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7.0 Acknowledgments

The Desalination and Water Purification Research and Development Program, Bureau of Reclamation, sponsored this research. Note participants and their credentials here. However, only those persons who made meaningful contributions to the research or to the report should be included.] We would like to acknowledge the hard work and dedication to this project from the following team members:

- Eric Dole PE, PSAP
- Robert E. Boysen, PE
- Jessica Johnson
- Tyler LeClear
- Kamyar Sardari, PhD, PE
- Jeremy Beard, PE
- Guillermo Delgado, PhD. PE
- Mike Mickley, PhD. PE

Metric Conversions (optional but highly recommended)

Provide metric equivalents for non-metric units used in the text:

Unit	Metric equivalent
1 gallon	3.785 liters
1 gallon per minute	3.785 liters per minute
1 gallon per square foot of membrane area per day	40.74 liters per square meter per day
1 inch	2.54 centimeters
1 million gallons per day	3,785 cubic meters per day
1 pound per square inch	6.895 kilopascals
1 square foot	0.093 square meters
°F (temperature measurement)	(°F–32) × 0.556 = °C
1 °F (temperature change or difference)	0.556 °C

Appendix A

A-1 – Project Sponsors and Approximate Donation Amounts

Special thanks to our industry partners who donated over $480,000 of in-kind equipment and labor!

- *RRCC* – donated ~$80,000 to the project including the pilot trailer, students to help operate it, and their electrician to install new power supply and distribution.

- *Powell Electro-coagulation (EC*) – coordinated the donation of a $150,000, 1.5 gpm EC unit from an anonymous donor to RRCC.

- *Salt Miner Inc* – waived the rental fee of $86,400 for a 1 gph concentrate treatment vacuum assisted electro-distillation (VAED) system for the 8-month duration of the field testing.

- *Endress+Hauser* - donated ~$50,000 for five pico magmeters, seven inline pH, temperature, ORP, conductivity and turbidity monitors, and one inline total hardness monitor.

- *Rockwell Automation* – donated ~ $7,640 for four 120 V power quality monitors to help monitor energy intensity of each unit process being evaluated and one VFD to operate the RW hose pump.

- *Flowrox* – donated $7,700 for the 1.5 HP raw water hose pump so we can easily and accurately dose the process train with a consistent raw water flow rate.

- *Blue-White* – donated three automated peristaltic chemical feed skids worth $28,611.

- *Garney Construction* – donated ~$2,000 in miscellaneous pipe and materials used to build the clarifier.

- *Lyons Filter* – donated ~$4,300 of labor and materials (repurposed CST and miscellaneous tanks used throughout the process) and agreed to help assemble trailer instrumentation and automation.

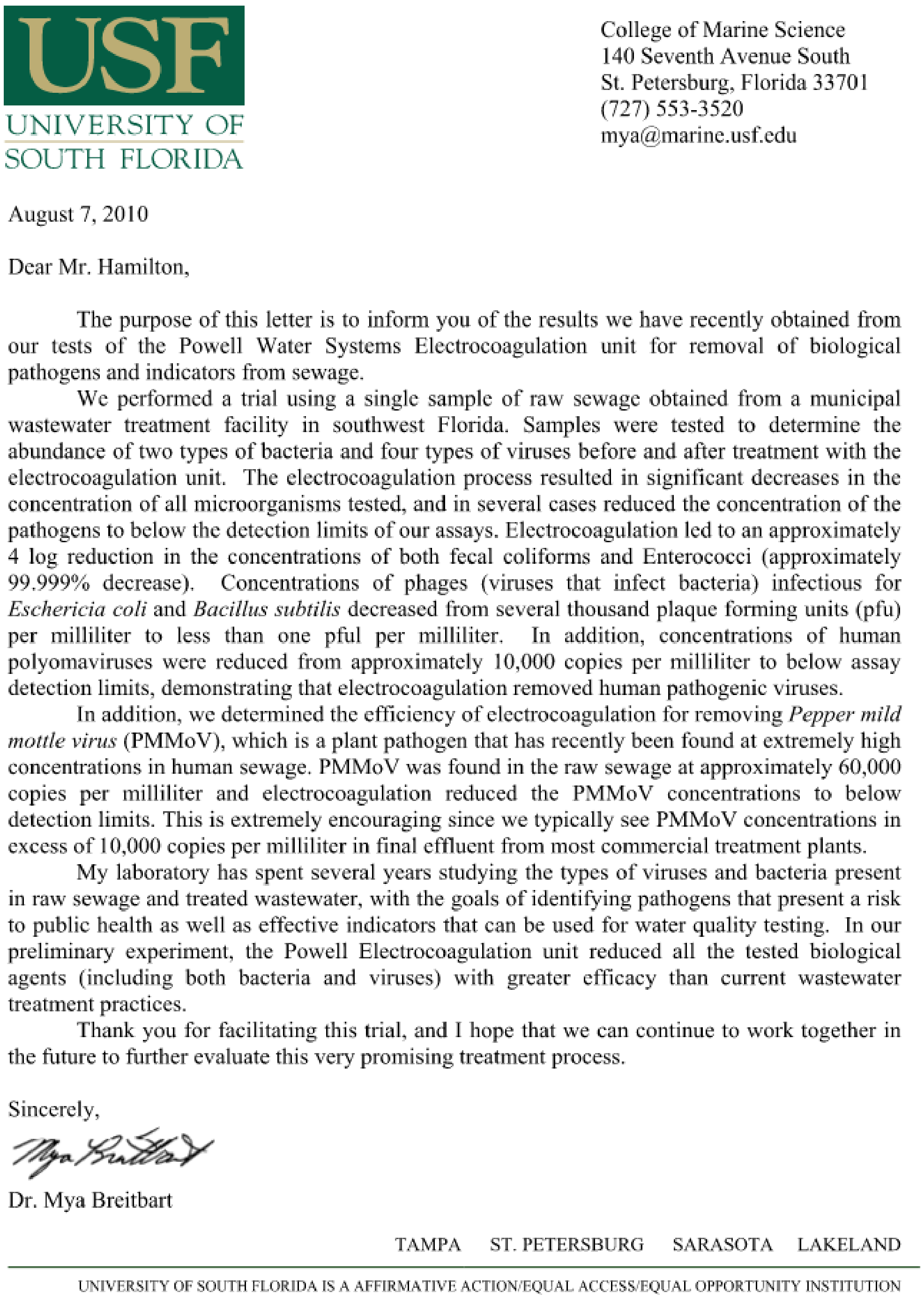
- *Harrington Plastics* – donated $3,770 in misc. PVC pipe, fittings, and valves.

- *Mickley & Associates* – reduced hourly billing rate by 50% to provide QA/QC services valued at $9,640.

- *Garver LLC* – donated $16,236 of our QA/QC time to the project per our proposal.

- *Colorado Analytical* – reduced unit sample costs for all analytes by 10%.

A-2 – University of South Florida Study on EC Pathogen and Virus Kill Efficiency



USF
UNIVERSITY OF
SOUTH FLORIDA

College of Marine Science
140 Seventh Avenue South
St. Petersburg, Florida 33701
(727) 553-3520
mya@marine.usf.edu

August 7, 2010

Dear Mr. Hamilton,

The purpose of this letter is to inform you of the results we have recently obtained from our tests of the Powell Water Systems Electrocoagulation unit for removal of biological pathogens and indicators from sewage.

We performed a trial using a single sample of raw sewage obtained from a municipal wastewater treatment facility in southwest Florida. Samples were tested to determine the abundance of two types of bacteria and four types of viruses before and after treatment with the electrocoagulation unit. The electrocoagulation process resulted in significant decreases in the concentration of all microorganisms tested, and in several cases reduced the concentration of the pathogens to below the detection limits of our assays. Electrocoagulation led to an approximately 4 log reduction in the concentrations of both fecal coliforms and Enterococci (approximately 99.999% decrease). Concentrations of phages (viruses that infect bacteria) infectious for *Eschericia coli* and *Bacillus subtilis* decreased from several thousand plaque forming units (pfu) per milliliter to less than one pful per milliliter. In addition, concentrations of human polyomaviruses were reduced from approximately 10,000 copies per milliliter to below assay detection limits, demonstrating that electrocoagulation removed human pathogenic viruses.

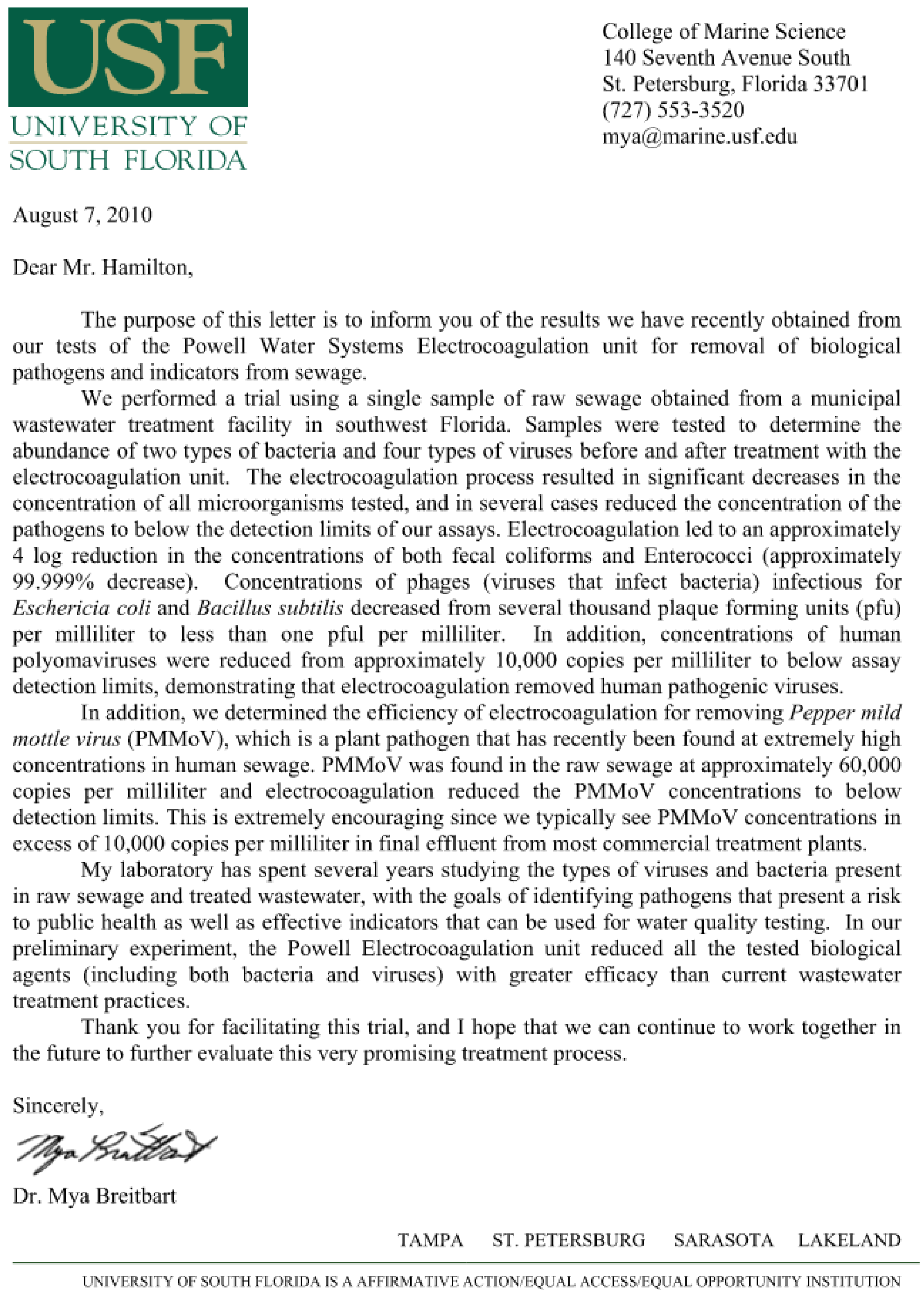
In addition, we determined the efficiency of electrocoagulation for removing *Pepper mild mottle virus* (PMMoV), which is a plant pathogen that has recently been found at extremely high concentrations in human sewage. PMMoV was found in the raw sewage at approximately 60,000 copies per milliliter and electrocoagulation reduced the PMMoV concentrations to below detection limits. This is extremely encouraging since we typically see PMMoV concentrations in excess of 10,000 copies per milliliter in final effluent from most commercial treatment plants.

My laboratory has spent several years studying the types of viruses and bacteria present in raw sewage and treated wastewater, with the goals of identifying pathogens that present a risk to public health as well as effective indicators that can be used for water quality testing. In our preliminary experiment, the Powell Electrocoagulation unit reduced all the tested biological agents (including both bacteria and viruses) with greater efficacy than current wastewater treatment practices.

Thank you for facilitating this trial, and I hope that we can continue to work together in the future to further evaluate this very promising treatment process.

Sincerely,

Dr. Mya Breitbart



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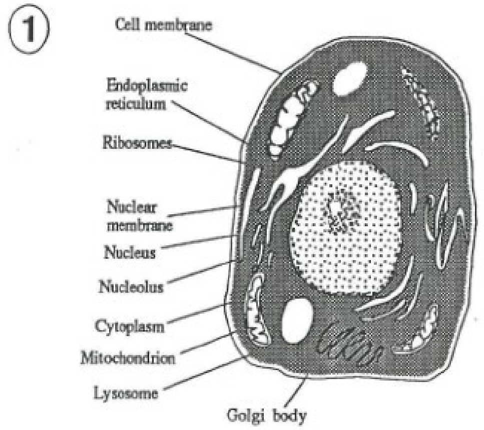
Sincerely,

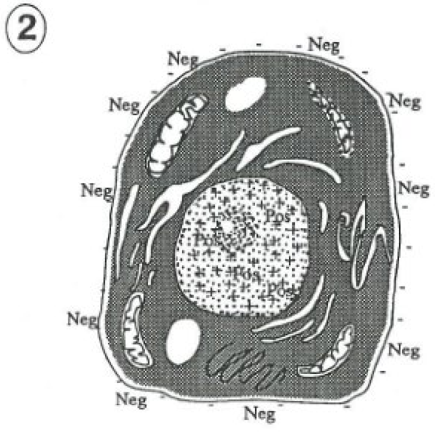
Dr. Mya Breitbart

TAMPA ST. PETERSBURG SARASOTA LAKELAND

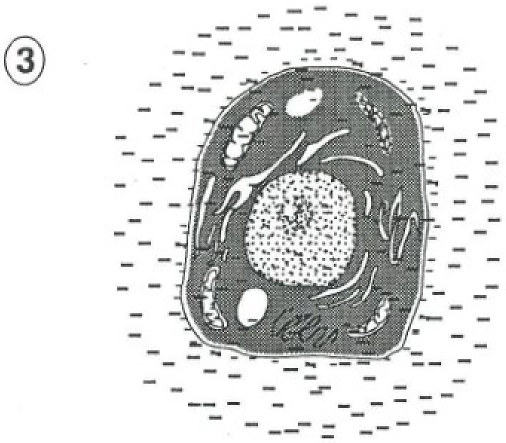
ELECTROPHORETIC / ELECTROCIDAL- EFFECT ON NUCLEATED MICROORGANISMS " CHEMICAL FREE "



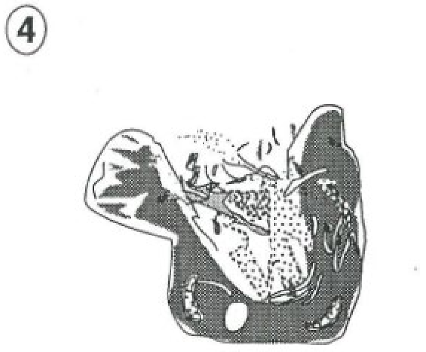
NORMAL NUCLEATED MICROORGANISM



TYPICAL ELECTRICAL NET SURFACE CHARGE ON OUTER MEMBRANE



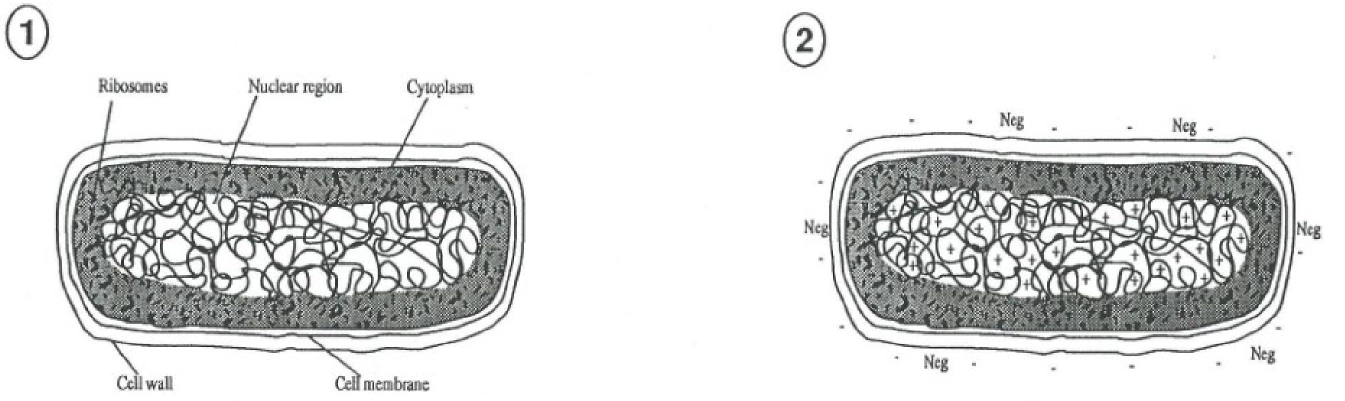
INDUCED PLASMA SURCHARGE ON OUTER MEMBRANE



CAVITATION AND DNA DESTRUCTION (DEATH OF THE MICROORGANISM)

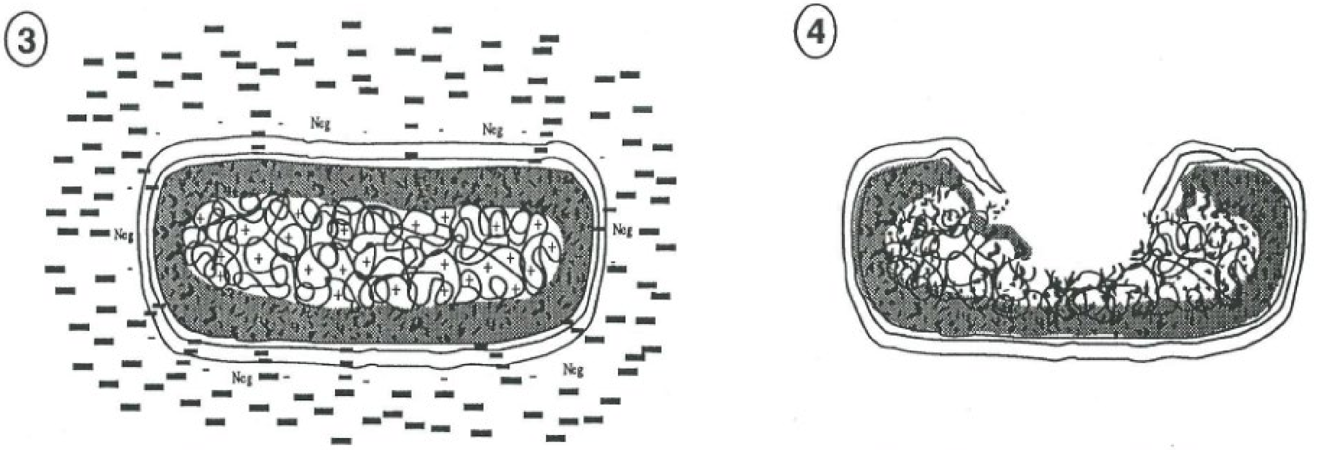
Courtesy: Russell Renk, PhD, PE – Western Research Institute

ELECTROPHORETIC / ELECTROCIDAL- EFFECT ON NON- NUCLEATED MICROORGANISMS "CHEMICAL FREE"



NORMAL NON-NUCLEATED MICROORGANISM

TYPICAL ELECTRICAL NET SURFACE CHARGE ON OUTER MEMBRANE

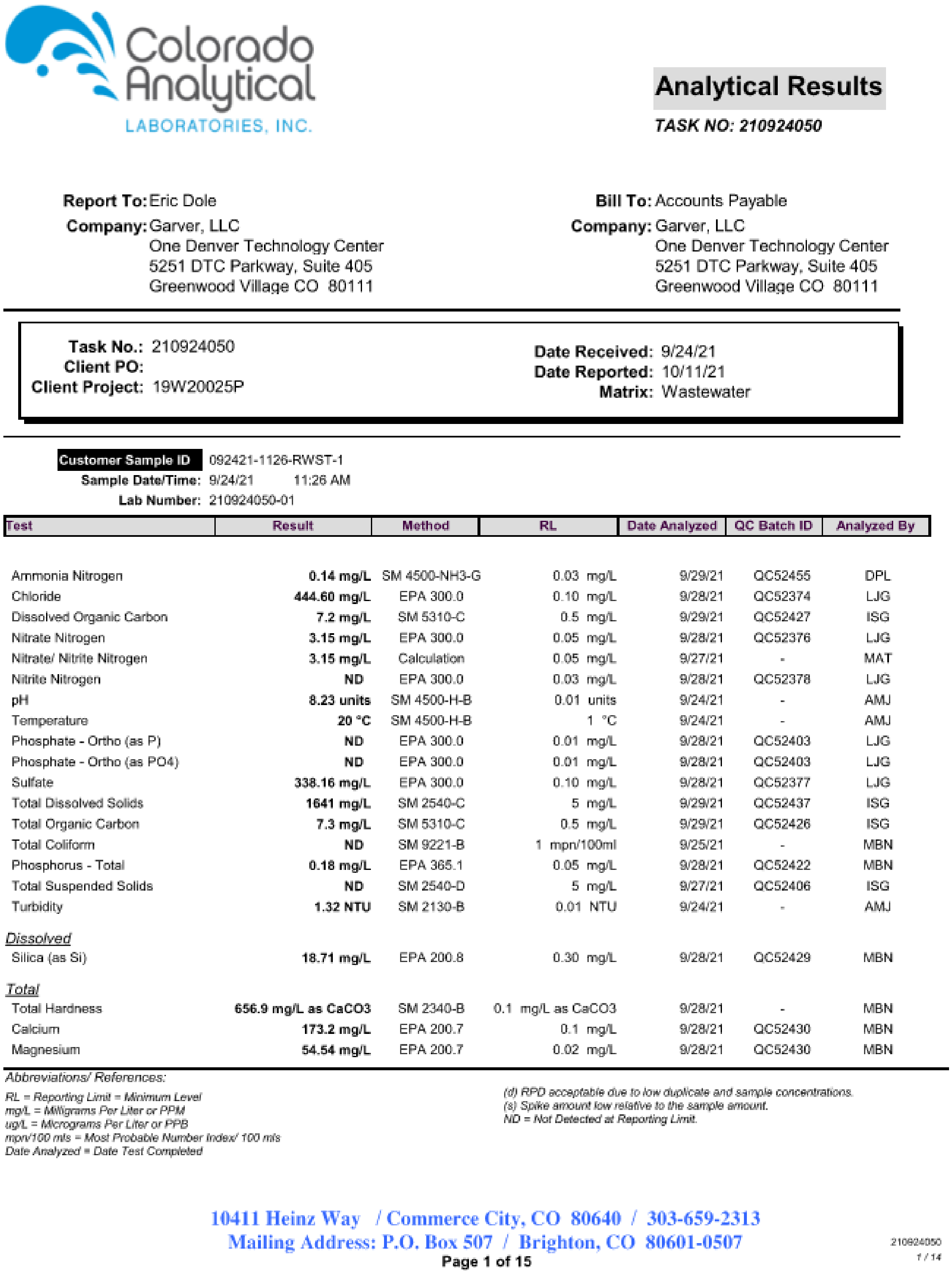


INDUCED PLASMA SURCHARGE ON OUTER MEMBRANE

CAVITATION AND DNA DESTRUCTION (DEATH OF THE MICROORGANISM)

Courtesy: Russell Renk, PhD, PE – Western Research Institute

A-3 – Colorado Analytical Sample Lab Results of Target Analytes



Colorado Analytical
LABORATORIES, INC.

Analytical Results

TASK NO: 210924050

Report To: Eric Dole
Company: Garver, LLC
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village CO 80111

Bill To: Accounts Payable
Company: Garver, LLC
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village CO 80111

Task No.: 210924050
Client PO:
Client Project: 19W20025P

Date Received: 9/24/21
Date Reported: 10/11/21
Matrix: Wastewater

Customer Sample ID 092421-1126-RWST-1
Sample Date/Time: 9/24/21 11:26 AM
Lab Number: 210924050-01

Test	Result	Method	RL	Date Analyzed	QC Batch ID	Analyzed By
Ammonia Nitrogen	**0.14 mg/L**	SM 4500-NH3-G	0.03 mg/L	9/29/21	QC52455	DPL
Chloride	**444.60 mg/L**	EPA 300.0	0.10 mg/L	9/28/21	QC52374	LJG
Dissolved Organic Carbon	**7.2 mg/L**	SM 5310-C	0.5 mg/L	9/29/21	QC52427	ISG
Nitrate Nitrogen	**3.15 mg/L**	EPA 300.0	0.05 mg/L	9/28/21	QC52376	LJG
Nitrate/ Nitrite Nitrogen	**3.15 mg/L**	Calculation	0.05 mg/L	9/27/21	-	MAT
Nitrite Nitrogen	**ND**	EPA 300.0	0.03 mg/L	9/28/21	QC52378	LJG
pH	**8.23 units**	SM 4500-H-B	0.01 units	9/24/21	-	AMJ
Temperature	**20 °C**	SM 4500-H-B	1 °C	9/24/21	-	AMJ
Phosphate - Ortho (as P)	**ND**	EPA 300.0	0.01 mg/L	9/28/21	QC52403	LJG
Phosphate - Ortho (as PO4)	**ND**	EPA 300.0	0.01 mg/L	9/28/21	QC52403	LJG
Sulfate	**338.16 mg/L**	EPA 300.0	0.10 mg/L	9/28/21	QC52377	LJG
Total Dissolved Solids	**1641 mg/L**	SM 2540-C	5 mg/L	9/29/21	QC52437	ISG
Total Organic Carbon	**7.3 mg/L**	SM 5310-C	0.5 mg/L	9/29/21	QC52426	ISG
Total Coliform	**ND**	SM 9221-B	1 mpn/100ml	9/25/21	-	MBN
Phosphorus - Total	**0.18 mg/L**	EPA 365.1	0.05 mg/L	9/28/21	QC52422	MBN
Total Suspended Solids	**ND**	SM 2540-D	5 mg/L	9/27/21	QC52406	ISG
Turbidity	**1.32 NTU**	SM 2130-B	0.01 NTU	9/24/21	-	AMJ
Dissolved						
Silica (as Si)	**18.71 mg/L**	EPA 200.8	0.30 mg/L	9/28/21	QC52429	MBN
Total						
Total Hardness	**656.9 mg/L as CaCO3**	SM 2340-B	0.1 mg/L as CaCO3	9/28/21	-	MBN
Calcium	**173.2 mg/L**	EPA 200.7	0.1 mg/L	9/28/21	QC52430	MBN
Magnesium	**54.54 mg/L**	EPA 200.7	0.02 mg/L	9/28/21	QC52430	MBN

Abbreviations/ References:

RL = Reporting Limit = Minimum Level
mg/L = Milligrams Per Liter or PPM
ug/L = Micrograms Per Liter or PPB
mpn/100 mls = Most Probable Number Index/ 100 mls
Date Analyzed = Date Test Completed

(d) RPD acceptable due to low duplicate and sample concentrations.
(s) Spike amount low relative to the sample amount.
ND = Not Detected at Reporting Limit.

10411 Heinz Way / Commerce City, CO 80640 / 303-659-2313
Mailing Address: P.O. Box 507 / Brighton, CO 80601-0507
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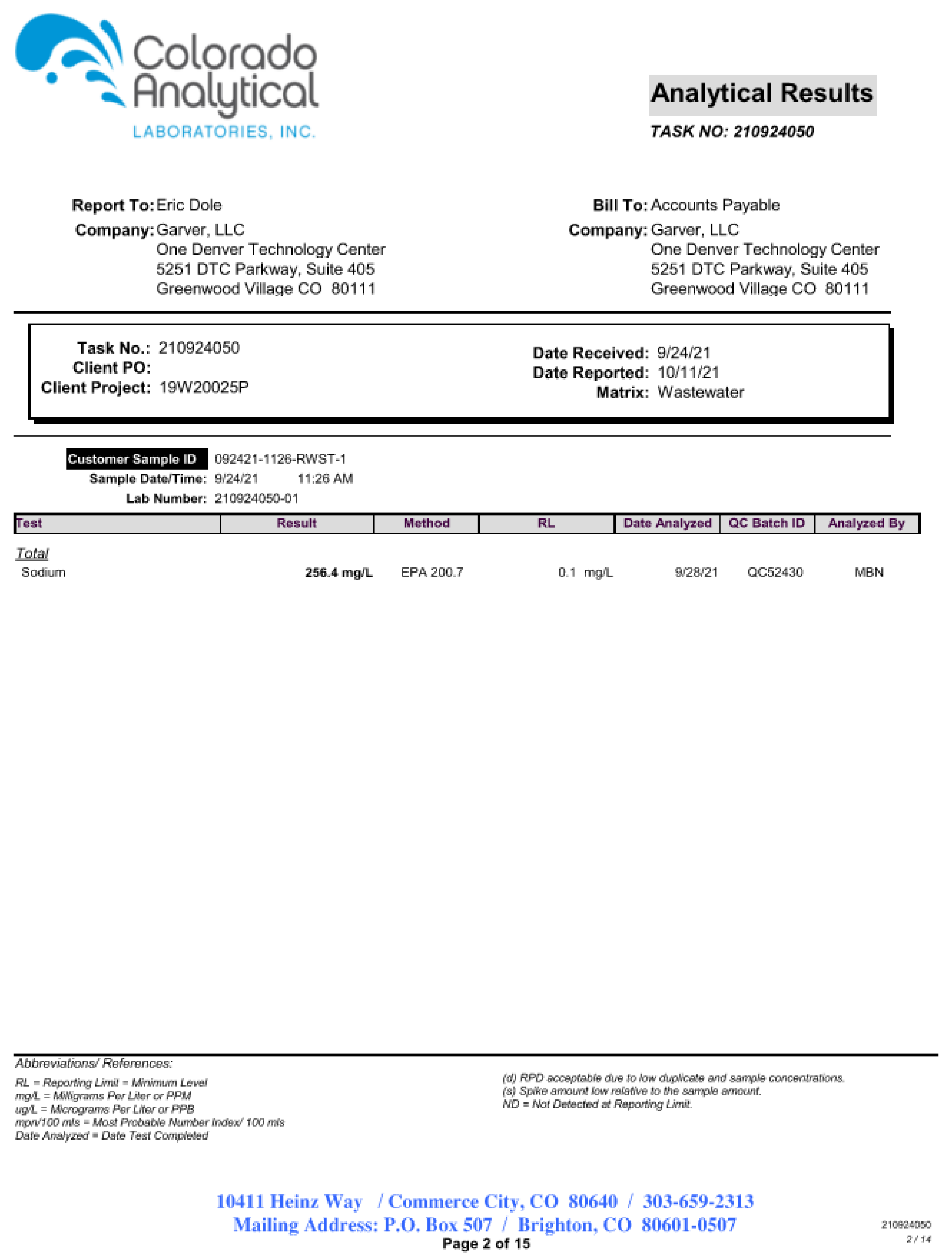


Colorado Analytical
LABORATORIES, INC.

Analytical Results

TASK NO: 210924050

Report To: Eric Dole
Company: Garver, LLC
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village CO 80111

Bill To: Accounts Payable
Company: Garver, LLC
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village CO 80111

Task No.: 210924050
Client PO:
Client Project: 19W20025P

Date Received: 9/24/21
Date Reported: 10/11/21
Matrix: Wastewater

Customer Sample ID 092421-1126-RWST-1
Sample Date/Time: 9/24/21 11:26 AM
Lab Number: 210924050-01

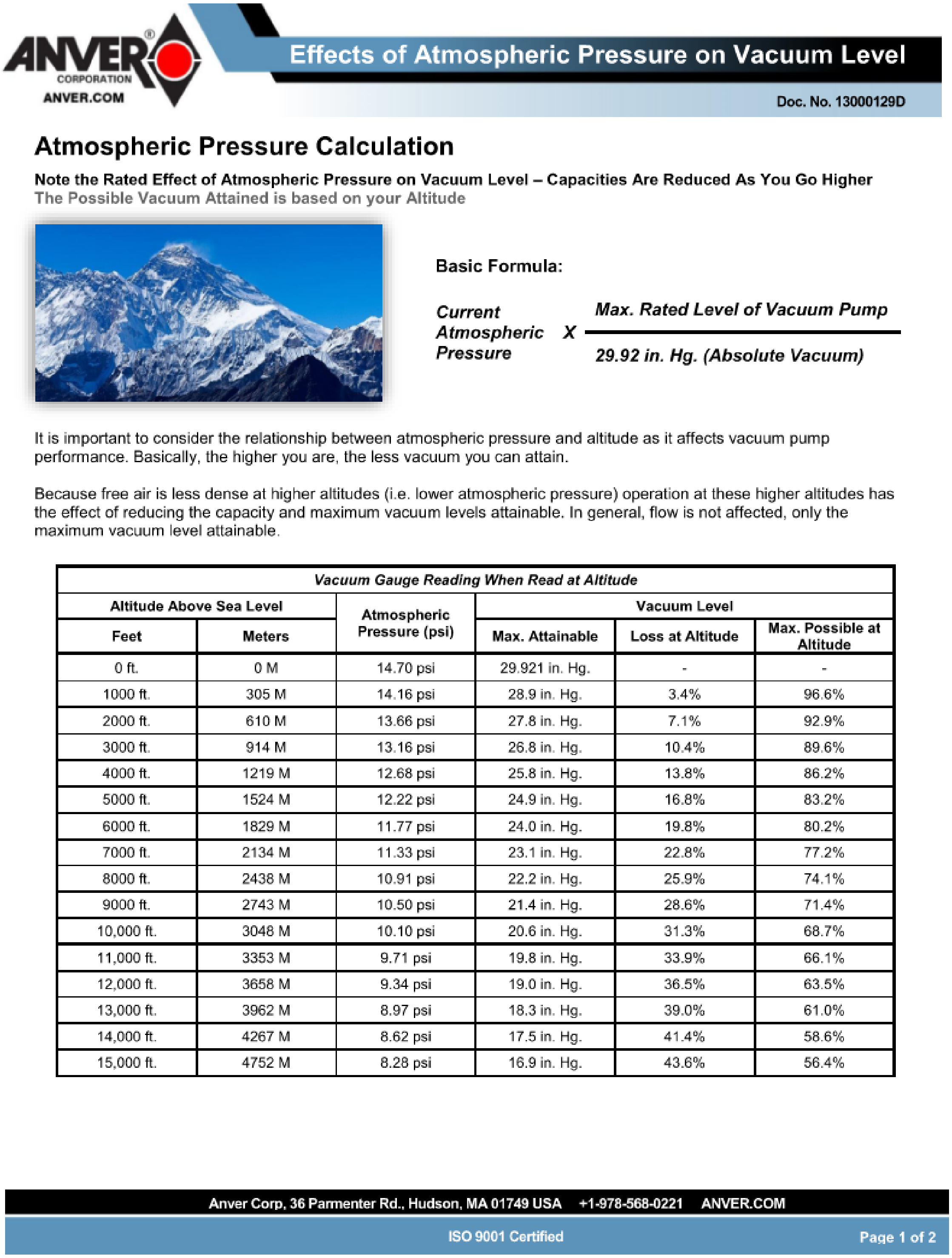
Test	Result	Method	RL	Date Analyzed	QC Batch ID	Analyzed By
Total						
Sodium	**256.4 mg/L**	EPA 200.7	0.1 mg/L	9/28/21	QC52430	MBN

Abbreviations/ References:

RL = Reporting Limit = Minimum Level
mg/L = Milligrams Per Liter or PPM
ug/L = Micrograms Per Liter or PPB
mpn/100 mls = Most Probable Number Index/ 100 mls
Date Analyzed = Date Test Completed

(d) RPD acceptable due to low duplicate and sample concentrations.
(s) Spike amount low relative to the sample amount.
ND = Not Detected at Reporting Limit.

A-4 – Vacuum Tables as a Function of Elevation and Atmospheric Pressure



ANVER CORPORATION
ANVER.COM

Effects of Atmospheric Pressure on Vacuum Level

Doc. No. 13000129D

Atmospheric Pressure Calculation

Note the Rated Effect of Atmospheric Pressure on Vacuum Level – Capacities Are Reduced As You Go Higher
The Possible Vacuum Attained is based on your Altitude

Basic Formula:

$$\textit{Current Atmospheric Pressure} \times \frac{\textit{Max. Rated Level of Vacuum Pump}}{\textit{29.92 in. Hg. (Absolute Vacuum)}}$$

It is important to consider the relationship between atmospheric pressure and altitude as it affects vacuum pump performance. Basically, the higher you are, the less vacuum you can attain.

Because free air is less dense at higher altitudes (i.e. lower atmospheric pressure) operation at these higher altitudes has the effect of reducing the capacity and maximum vacuum levels attainable. In general, flow is not affected, only the maximum vacuum level attainable.

Vacuum Gauge Reading When Read at Altitude					
Altitude Above Sea Level		**Atmospheric Pressure (psi)**	**Vacuum Level**		
Feet	**Meters**		**Max. Attainable**	**Loss at Altitude**	**Max. Possible at Altitude**
0 ft.	0 M	14.70 psi	29.921 in. Hg.	-	-
1000 ft.	305 M	14.16 psi	28.9 in. Hg.	3.4%	96.6%
2000 ft.	610 M	13.66 psi	27.8 in. Hg.	7.1%	92.9%
3000 ft.	914 M	13.16 psi	26.8 in. Hg.	10.4%	89.6%
4000 ft.	1219 M	12.68 psi	25.8 in. Hg.	13.8%	86.2%
5000 ft.	1524 M	12.22 psi	24.9 in. Hg.	16.8%	83.2%
6000 ft.	1829 M	11.77 psi	24.0 in. Hg.	19.8%	80.2%
7000 ft.	2134 M	11.33 psi	23.1 in. Hg.	22.8%	77.2%
8000 ft.	2438 M	10.91 psi	22.2 in. Hg.	25.9%	74.1%
9000 ft.	2743 M	10.50 psi	21.4 in. Hg.	28.6%	71.4%
10,000 ft.	3048 M	10.10 psi	20.6 in. Hg.	31.3%	68.7%
11,000 ft.	3353 M	9.71 psi	19.8 in. Hg.	33.9%	66.1%
12,000 ft.	3658 M	9.34 psi	19.0 in. Hg.	36.5%	63.5%
13,000 ft.	3962 M	8.97 psi	18.3 in. Hg.	39.0%	61.0%
14,000 ft.	4267 M	8.62 psi	17.5 in. Hg.	41.4%	58.6%
15,000 ft.	4752 M	8.28 psi	16.9 in. Hg.	43.6%	56.4%

Anver Corp, 36 Parmenter Rd., Hudson, MA 01749 USA +1-978-568-0221 ANVER.COM

ISO 9001 Certified

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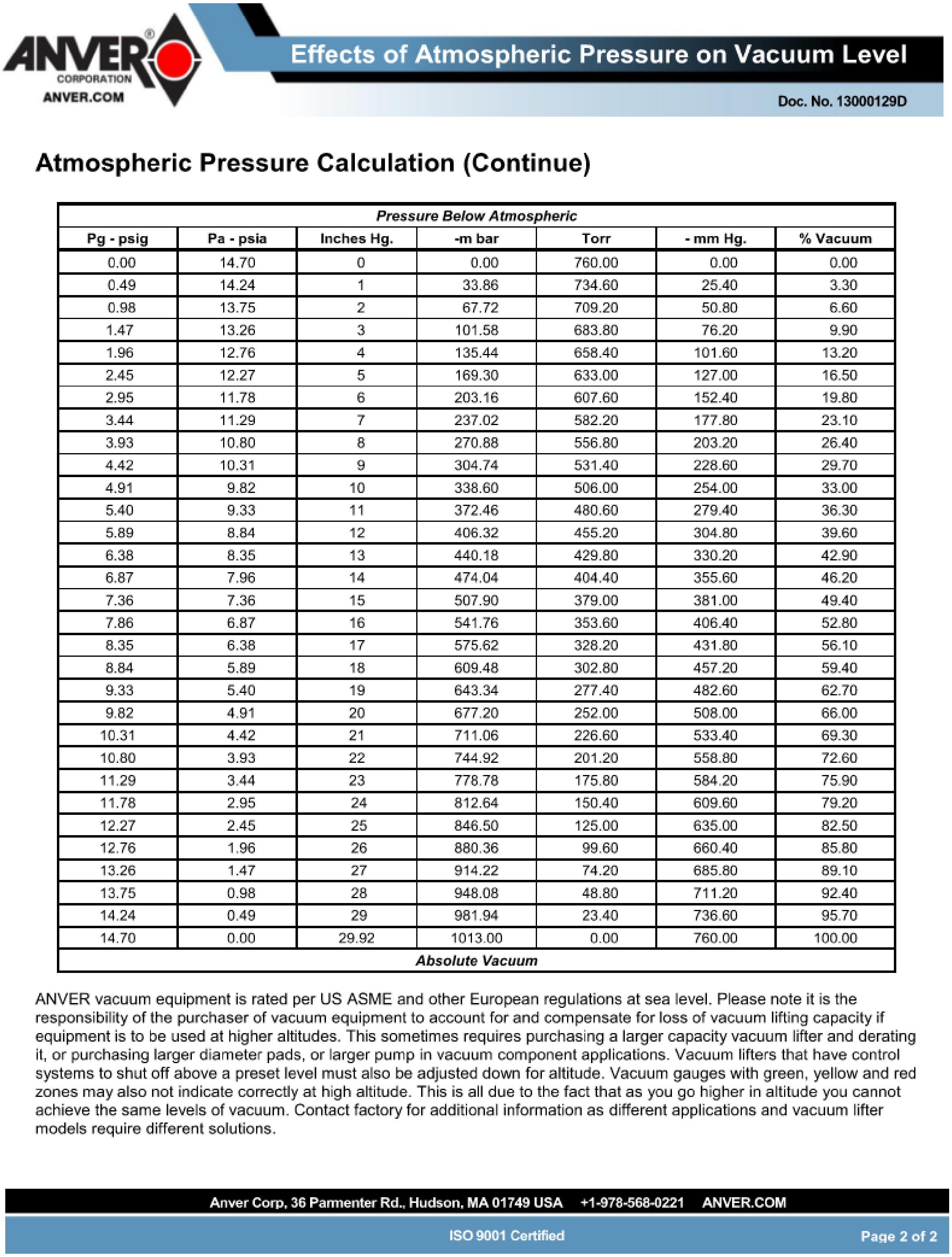

ANVER® CORPORATION
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Effects of Atmospheric Pressure on Vacuum Level

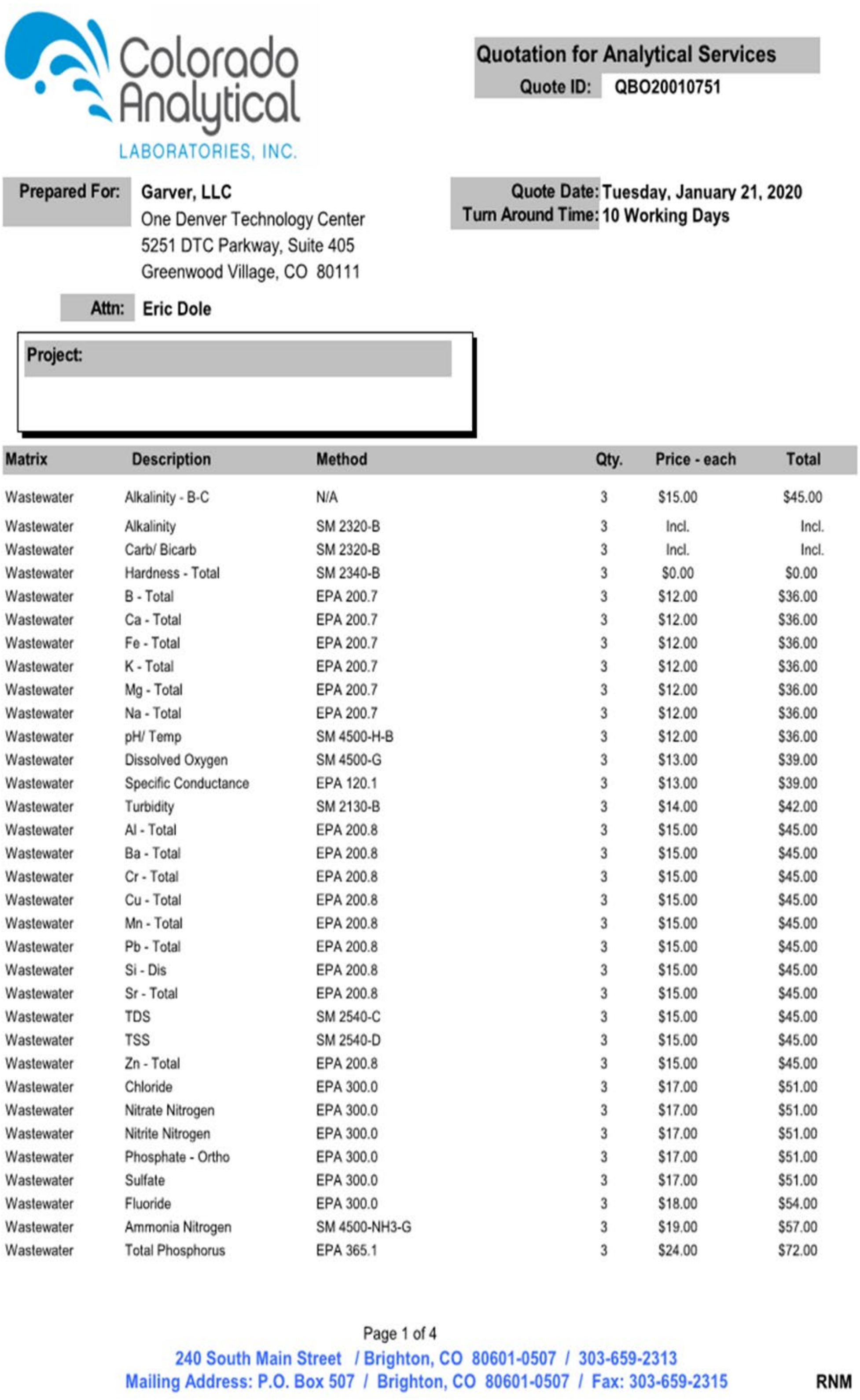
Doc. No. 13000129D

Atmospheric Pressure Calculation (Continue)

Pressure Below Atmospheric						
Pg - psig	**Pa - psia**	**Inches Hg.**	**-m bar**	**Torr**	**- mm Hg.**	**% Vacuum**
0.00	14.70	0	0.00	760.00	0.00	0.00
0.49	14.24	1	33.86	734.60	25.40	3.30
0.98	13.75	2	67.72	709.20	50.80	6.60
1.47	13.26	3	101.58	683.80	76.20	9.90
1.96	12.76	4	135.44	658.40	101.60	13.20
2.45	12.27	5	169.30	633.00	127.00	16.50
2.95	11.78	6	203.16	607.60	152.40	19.80
3.44	11.29	7	237.02	582.20	177.80	23.10
3.93	10.80	8	270.88	556.80	203.20	26.40
4.42	10.31	9	304.74	531.40	228.60	29.70
4.91	9.82	10	338.60	506.00	254.00	33.00
5.40	9.33	11	372.46	480.60	279.40	36.30
5.89	8.84	12	406.32	455.20	304.80	39.60
6.38	8.35	13	440.18	429.80	330.20	42.90
6.87	7.96	14	474.04	404.40	355.60	46.20
7.36	7.36	15	507.90	379.00	381.00	49.40
7.86	6.87	16	541.76	353.60	406.40	52.80
8.35	6.38	17	575.62	328.20	431.80	56.10
8.84	5.89	18	609.48	302.80	457.20	59.40
9.33	5.40	19	643.34	277.40	482.60	62.70
9.82	4.91	20	677.20	252.00	508.00	66.00
10.31	4.42	21	711.06	226.60	533.40	69.30
10.80	3.93	22	744.92	201.20	558.80	72.60
11.29	3.44	23	778.78	175.80	584.20	75.90
11.78	2.95	24	812.64	150.40	609.60	79.20
12.27	2.45	25	846.50	125.00	635.00	82.50
12.76	1.96	26	880.36	99.60	660.40	85.80
13.26	1.47	27	914.22	74.20	685.80	89.10
13.75	0.98	28	948.08	48.80	711.20	92.40
14.24	0.49	29	981.94	23.40	736.60	95.70
14.70	0.00	29.92	1013.00	0.00	760.00	100.00
Absolute Vacuum						

ANVER vacuum equipment is rated per US ASME and other European regulations at sea level. Please note it is the responsibility of the purchaser of vacuum equipment to account for and compensate for loss of vacuum lifting capacity if equipment is to be used at higher altitudes. This sometimes requires purchasing a larger capacity vacuum lifter and derating it, or purchasing larger diameter pads, or larger pump in vacuum component applications. Vacuum lifters that have control systems to shut off above a preset level must also be adjusted down for altitude. Vacuum gauges with green, yellow and red zones may also not indicate correctly at high altitude. This is all due to the fact that as you go higher in altitude you cannot achieve the same levels of vacuum. Contact factory for additional information as different applications and vacuum lifter models require different solutions.

A-5 – Colorado Analytical Quote for Water Quality Analysis



Colorado Analytical
LABORATORIES, INC.

Quotation for Analytical Services

Quote ID: QBO20010751

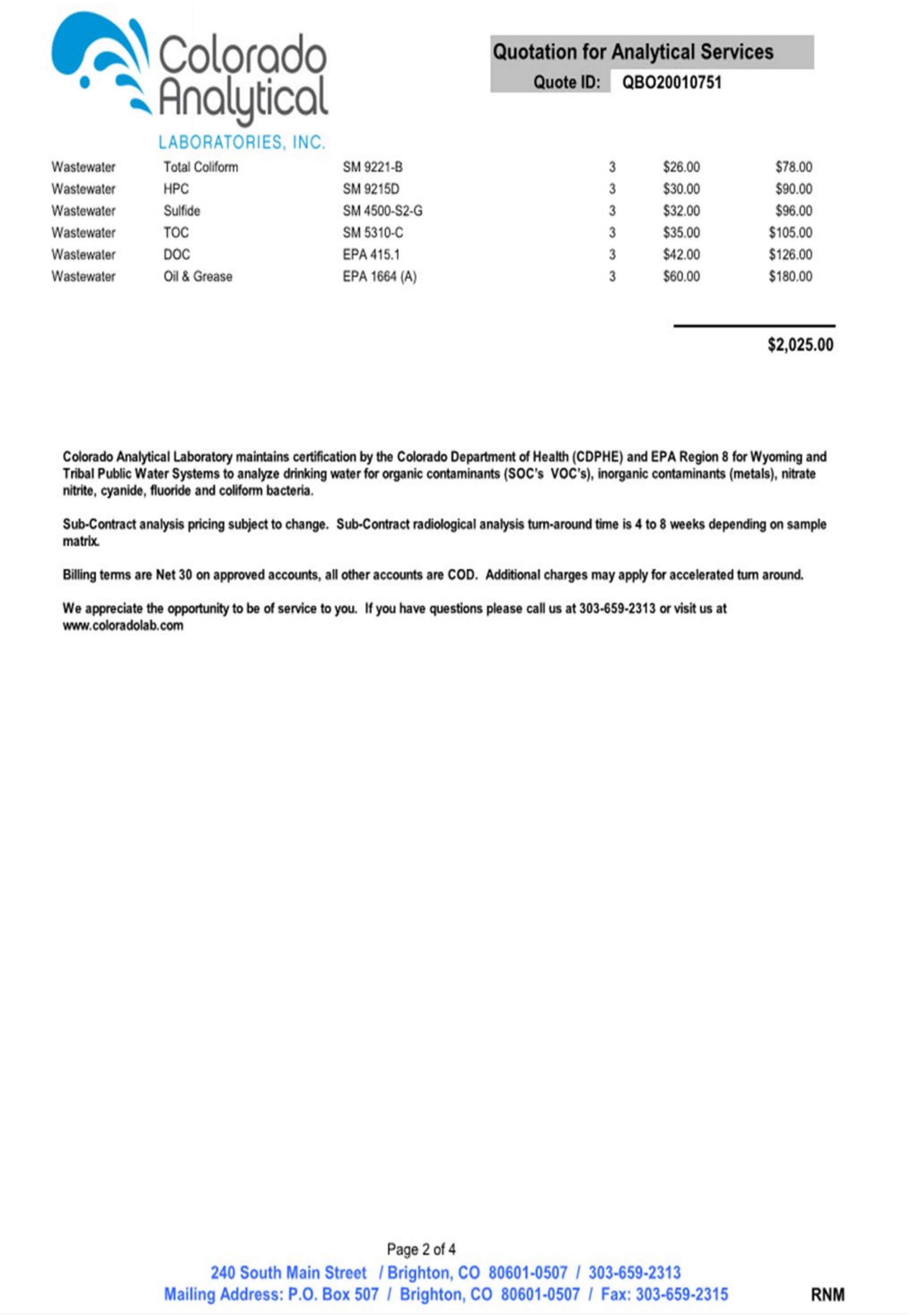
Prepared For: **Garver, LLC**
One Denver Technology Center
5251 DTC Parkway, Suite 405
Greenwood Village, CO 80111

Attn: **Eric Dole**

Quote Date: Tuesday, January 21, 2020
Turn Around Time: 10 Working Days

Project:

Matrix	Description	Method	Qty.	Price - each	Total
Wastewater	Alkalinity - B-C	N/A	3	$15.00	$45.00
Wastewater	Alkalinity	SM 2320-B	3	Incl.	Incl.
Wastewater	Carb/ Bicarb	SM 2320-B	3	Incl.	Incl.
Wastewater	Hardness - Total	SM 2340-B	3	$0.00	$0.00
Wastewater	B - Total	EPA 200.7	3	$12.00	$36.00
Wastewater	Ca - Total	EPA 200.7	3	$12.00	$36.00
Wastewater	Fe - Total	EPA 200.7	3	$12.00	$36.00
Wastewater	K - Total	EPA 200.7	3	$12.00	$36.00
Wastewater	Mg - Total	EPA 200.7	3	$12.00	$36.00
Wastewater	Na - Total	EPA 200.7	3	$12.00	$36.00
Wastewater	pH/ Temp	SM 4500-H-B	3	$12.00	$36.00
Wastewater	Dissolved Oxygen	SM 4500-G	3	$13.00	$39.00
Wastewater	Specific Conductance	EPA 120.1	3	$13.00	$39.00
Wastewater	Turbidity	SM 2130-B	3	$14.00	$42.00
Wastewater	Al - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	Ba - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	Cr - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	Cu - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	Mn - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	Pb - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	Si - Dis	EPA 200.8	3	$15.00	$45.00
Wastewater	Sr - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	TDS	SM 2540-C	3	$15.00	$45.00
Wastewater	TSS	SM 2540-D	3	$15.00	$45.00
Wastewater	Zn - Total	EPA 200.8	3	$15.00	$45.00
Wastewater	Chloride	EPA 300.0	3	$17.00	$51.00
Wastewater	Nitrate Nitrogen	EPA 300.0	3	$17.00	$51.00
Wastewater	Nitrite Nitrogen	EPA 300.0	3	$17.00	$51.00
Wastewater	Phosphate - Ortho	EPA 300.0	3	$17.00	$51.00
Wastewater	Sulfate	EPA 300.0	3	$17.00	$51.00
Wastewater	Fluoride	EPA 300.0	3	$18.00	$54.00
Wastewater	Ammonia Nitrogen	SM 4500-NH3-G	3	$19.00	$57.00
Wastewater	Total Phosphorus	EPA 365.1	3	$24.00	$72.00



Colorado Analytical LABORATORIES, INC.

Quotation for Analytical Services

Quote ID: QBO20010751

Wastewater	Total Coliform	SM 9221-B	3	$26.00	$78.00
Wastewater	HPC	SM 9215D	3	$30.00	$90.00
Wastewater	Sulfide	SM 4500-S2-G	3	$32.00	$96.00
Wastewater	TOC	SM 5310-C	3	$35.00	$105.00
Wastewater	DOC	EPA 415.1	3	$42.00	$126.00
Wastewater	Oil & Grease	EPA 1664 (A)	3	$60.00	$180.00
					$2,025.00

Colorado Analytical Laboratory maintains certification by the Colorado Department of Health (CDPHE) and EPA Region 8 for Wyoming and Tribal Public Water Systems to analyze drinking water for organic contaminants (SOC's VOC's), inorganic contaminants (metals), nitrate nitrite, cyanide, fluoride and coliform bacteria.

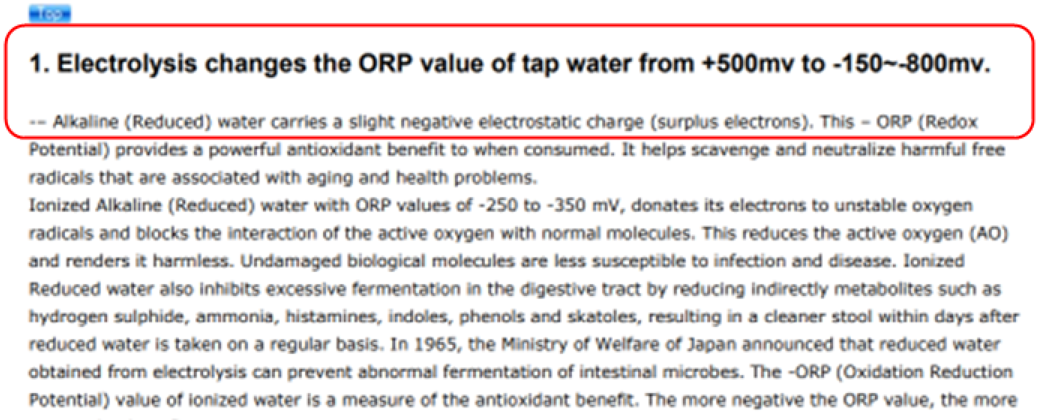
Sub-Contract analysis pricing subject to change. Sub-Contract radiological analysis turn-around time is 4 to 8 weeks depending on sample matrix.

Billing terms are Net 30 on approved accounts, all other accounts are COD. Additional charges may apply for accelerated turn around.

We appreciate the opportunity to be of service to you. If you have questions please call us at 303-659-2313 or visit us at www.coloradolab.com

A-6 – EC Antiscalent Impact on Membranes

EC *is similar to* electrolysis. Electrolysis causes water to be negatively charged (ie; - ORP)



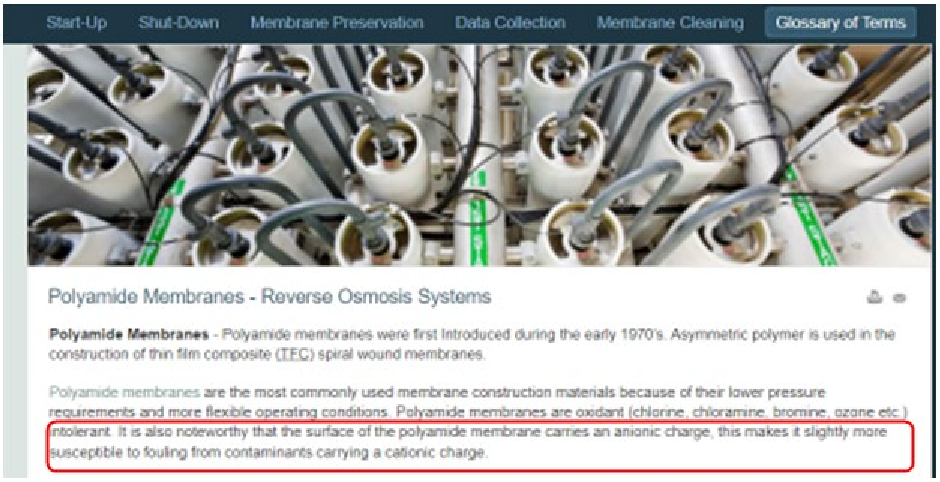
1. Electrolysis changes the ORP value of tap water from +500mv to -150~-800mv.

-- Alkaline (Reduced) water carries a slight negative electrostatic charge (surplus electrons). This – ORP (Redox Potential) provides a powerful antioxidant benefit to when consumed. It helps scavenge and neutralize harmful free radicals that are associated with aging and health problems.
Ionized Alkaline (Reduced) water with ORP values of -250 to -350 mV, donates its electrons to unstable oxygen radicals and blocks the interaction of the active oxygen with normal molecules. This reduces the active oxygen (AO) and renders it harmless. Undamaged biological molecules are less susceptible to infection and disease. Ionized Reduced water also inhibits excessive fermentation in the digestive tract by reducing indirectly metabolites such as hydrogen sulphide, ammonia, histamines, indoles, phenols and skatoles, resulting in a cleaner stool within days after reduced water is taken on a regular basis. In 1965, the Ministry of Welfare of Japan announced that reduced water obtained from electrolysis can prevent abnormal fermentation of intestinal microbes. The -ORP (Oxidation Reduction Potential) value of ionized water is a measure of the antioxidant benefit. The more negative the ORP value, the more scavenging benefit.

https://d11fdyfhxcs9cr.cloudfront.net/templates/517763/myimages/water_ionizer_and_its_function.pdf

Polyamide (PA) membranes are also negatively charged

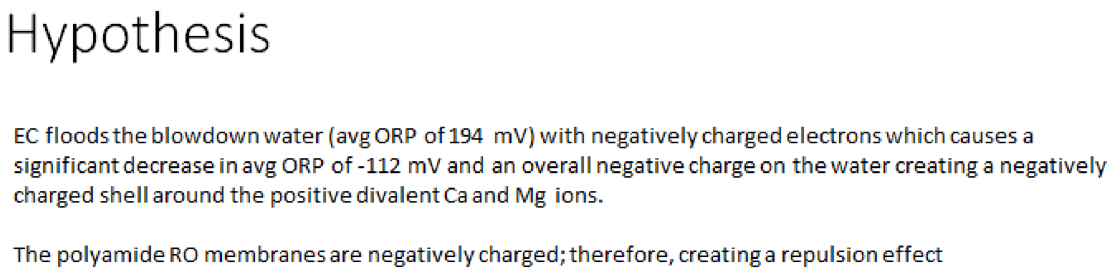
http://reverseosmosischemicals.com/reverse-osmosis-guides/reverse-osmosis-glossary-terms/polyamide-membranes-reverse-osmosis-systems



Polyamide Membranes - Reverse Osmosis Systems

Polyamide Membranes - Polyamide membranes were first Introduced during the early 1970's. Asymmetric polymer is used in the construction of thin film composite (TFC) spiral wound membranes.

Polyamide membranes are the most commonly used membrane construction materials because of their lower pressure requirements and more flexible operating conditions. Polyamide membranes are oxidant (chlorine, chloramine, bromine, ozone etc.) intolerant. It is also noteworthy that the surface of the polyamide membrane carries an anionic charge, this makes it slightly more susceptible to fouling from contaminants carrying a cationic charge.



Hypothesis

EC floods the blowdown water (avg ORP of 194 mV) with negatively charged electrons which causes a significant decrease in avg ORP of -112 mV and an overall negative charge on the water creating a negatively charged shell around the positive divalent Ca and Mg ions.

The polyamide RO membranes are negatively charged; therefore, creating a repulsion effect

A-7 – Benchtop Tests to Help Prove EC Contribution to Hardness Removal and "Electric Antiscalent" Hypothesis

By: Mike Mickley, PhD, PE | President Mickley & Associates

PROPOSED TESTS

- TEST 1: demonstrate visible and measurable effects of occurrence of scaling as a function of pH – synthetic solutions of RWST-1water
- TEST 2: repeat test 1 runs exposing the solutions to electric current to demonstrate the influence of electric current on precipitation kinetics
- TEST 3: demonstrate influence of electric current inhibition of scaling – synthetic solutions and bench-scale RO system

COMMENTS:

- General test descriptions here are guidelines; detailed test protocols need to be developed.

TEST 1 – OCCURRENCE OF SCALING AS A FUNCTION OF PH – synthetic solution

Test 1 GOAL = determine how much reduction of Ca is because of higher pH. TEST 1 will indicate the amount of Ca reduction as a function of pH. Test conditions will approximate trailer test conditions in terms of Ca levels and pH values as well as general composition. Total Ca removal is indicated by the trailer tests. Removal by pH will be indicated by the lab tests.

Test protocols

- A thorough literature search should be conducted to indicate kinetic and other factors involved in ion precipitation onset. Included in this will be indications of the correlation of turbidity with visible onset of precipitation.
- There are several variables (volumes, stir speed, amount of caustic to add, how often to take turbidity readings, etc.) that the run conditions for meaningful data need to be

determined by trial and error. The test protocols will be revised as needed until measurable and reproducible results are obtained.

- Test run times will start after caustic addition to the synthetic RWST-1 water.
- Tests will be run at different Ca concentrations

Desired test findings

- Whether onset precipitation effects can be seen visually
- Whether turbidity is a good indicator for path to equilibrium; how best to indicate equilibrium conditions
- The time to reach equilibrium
- How pH influences this time
- Reproducibility of results
- What happens when Ca, CO3 levels, and pH match the levels in RWST-1 and within the residence times encountered in the trailer before EC treatment and EC/RO treatment.

TEST 2 – DEMONSTRATION ELECTRIC CURRENT EFFECT ON SCALING - synthetic solution

Test 2 GOAL = determination of the effect of electric current on the scaling kinetics of feedwater to an RO system.

Test protocols

- The only difference between Test 1 and Test 2 is exposing the test solutions used in Test 1 to electric current.
- Exposure will be with a small EC system equipped with stainless steel (SS) electrodes and powering unit. The SS electrodes eliminate other EC effects taking place – other than flow of current.
- The voltage and amperage as well as exposure time will approximate that used in the original trailer test conditions.

Desired test findings

- The effect electrical current has on precipitation kinetics. This may be indicated by comparison of precipitation kinetics with and without exposure to electrical current.

TEST 3 – DEMONSTRATION ELECTRIC CURRENT EFFECT ON RO SCALING - synthetic solution

Test 3 GOAL = This test may not be needed to reach a conclusion – given the results of test 2. It would, however, be a more direct demonstration of the results indicated in test 2. The goal is to directly demonstrate the inhibition of scaling due to exposure of RO feed solution to electric current.

Test protocol

- The test runs will involve a small bench-scale RO system using flat sheet membranes.
- Run conditions will be determined by trial and error to define feed compositions that result in membrane scaling for solutions not exposed to electric current. Scaling may be indicated by reduced flux and changes in pressures.
- Once scaling feed compositions are defined, runs will be made exposing feed to electric current.

Desired finding

- Reduced RO scaling for feed solutions exposed to electric current.