

Crystal Clearwater Resources  
Final Technical Completion Report  
Produced Water Desalination  
Pilot Results



Prepared for  
NMPWRC



May 4<sup>th</sup>, 2022



September 15, 2022

Crystal Clearwater Resources  
Attn: Derek Pedersen  
2351 W. Northwest Highway #2320  
Dallas, TX 75220

RE: Final Technical Completion Report, Produced Water Desalination - Pilot Results

Dear Mr. Pedersen,

The New Mexico Produced Water Research Consortium (Consortium) is a government-university-industry partnership operated by New Mexico State University in collaboration with the New Mexico Environment Department. One of our roles is to coordinate and evaluate the testing of innovative produced water treatment technologies with the potential to safely and cost-effectively treat produced water to support fit-for-purpose uses of outside the oil and gas sector.

The Consortium established a project review team composed of technical members from the Consortium that included industry, academia, and non-government agency representatives. Those team members reviewed your test plan, visited the test site to observe operations, and have reviewed both your preliminary and final reports.

Our evaluation team found that the data and information provided in the final report accurately represents the overall results and performance of the pilot-testing conducted in cooperation with the Consortium.

We compliment you and your team on a successful pilot test, a well documented and thorough technical report, and a job well done. If we can be of any further assistance, please feel free to contact me at 505-859-1563 or by email at [mmhightower@q.com](mailto:mmhightower@q.com).

Sincerely,

A handwritten signature in black ink that reads "Mike Hightower". The signature is written in a cursive, slightly slanted style.

Mike Hightower, Program Director  
New Mexico Produced Water Research Consortium



Derek Pedersen  
2351 W Northwest Hwy #2320  
Dallas, TX, 75220

May 4<sup>th</sup>, 2022

Mike Hightower Program Director  
1780 E University Ave  
Las Cruces, New Mexico

[SUBJ: Produced Water Treatment Utilizing Compressor Station Waste-Heat](#)

Dear Mr. Hightower,

Crystal Clearwater Resources (CCR) is pleased to present the New Mexico Produced Water Research Consortium (NMPWRC) with the final report of a demonstration pilot that was completed in Q4 of 2021 at a compressor station located near the New Mexico-Texas border. This report summarizes all relevant information and data collected from the pilot on key findings, challenges encountered, and successes achieved as measured by key performance indicators.

The Low Temperature Distillation (LTDis<sup>®</sup>) technology was used successfully to demonstrate the ability to treat oilfield produced water at the compressor facility (the Border Demonstration). For this Demonstration, the D1 Pilot LTDis<sup>®</sup> Unit (D1) was driven by low grade exhaust waste heat from a Caterpillar 3608 compressor. Phase 1 of the D1 was completed solely for the oil & gas operator and its partners while Phase 2 of the D1 was completed jointly with an oil & gas operator, and the NMPWRC. CCR is able to share the data from both Phase 1 and 2 in this report.

The pilot tests successfully demonstrated the D1's ability to treat oilfield produced water with a Total Dissolved Solids (TDS) ranging from 100,000-170,000 ppm, for at least fifteen 24-hour operating days, while producing a distillate stream of less than 500 ppm TDS. This was achieved using only waste heat from a single gas compressor. Plant performance proved to be consistent throughout the range of water salinities received during the D1.

The D1 results and lessons learned from Phases 1 and 2 have provided sufficient data to support the successful design and deployment of a full-scale application of the LTDis<sup>®</sup> Technology at multiple water treatment locations throughout the Country.

CCR appreciates the opportunity to work with the New Mexico Produced Water Research Consortium to evaluate our innovative solution and looks forward to working on the commercial scale solution.

Best Regards,

A handwritten signature in blue ink, appearing to read "Derek Pedersen", with a long horizontal flourish extending to the right.

Derek Pedersen  
Co-Founder and CEO  
Crystal Clearwater Resources

## Disclaimer

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Any questions concerning the information, or its interpretation should be directed to:

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COO

Crystal Clearwater Resources (CCR)

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## Acknowledgments

Crystal Clearwater Resources (CCR) is pleased to present the New Mexico Produced Water Research Consortium (NMPWRC) with the final report of the pilot (D1) that was completed in Q4 of 2021 along the New Mexico-Texas border.

CCR appreciates the opportunity that the NMPWRC provided with this grant. The demonstration was a valuable showcase of our technology and was visited by members representing various companies such as Chevron Corporation, Occidental Petroleum (Oxy), Berry Petroleum and Pioneer Natural Resources along with members of the NMPWRC consortium.

We trust that this report meets the needs and expectations of the NMPWRC. If you have any questions, please contact CCR at any time.

## Informative Abstract

Crystal Clearwater Resources (CCR) was founded to address the treatment of challenging waters that are generated across multiple industries. Availability of freshwater is becoming scarce, and regulations are making it more difficult to dispose of contaminated wastewater. Our Low Temperature Distillation (LTD<sup>is</sup>) technology is a step-change solution that offers the flexibility to use traditional, as well as environmentally friendly renewable sources of thermal energy, in the purification of contaminated waters.

The technology pilot was conducted to demonstrate CCR's patented LTD<sup>is</sup> Technology and its ability to treat high saline produced waters while utilizing only waste heat captured from a compressor station. The purpose of the demonstration was to demonstrate that LTD<sup>is</sup> can effectively and efficiently increase water recovery and decrease the amount of brine required to be disposed, while producing a clean distillate.

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## 1 Executive Summary

The goal as outlined in the creation of the NMPWRC is to accelerate technology and process research, development, and implementation for environmentally sound, safe, and cost-effective reuse of produced water. In accordance with this goal, the purpose of the pilot testing completed by CCR was to demonstrate the ability of the LTDIs<sup>®</sup> technology to treat oilfield produced water with multi-stage distillation using only waste heat as a thermal energy source. The distillate produced was required to be of acceptable quality for industrial or non-potable reuse. The water quality data collected during the D1 was collected and is currently being studied and trialed for beneficial reuse in agriculture.

The Key Performance Objectives (KPIs) identified are:

- 1) **Waste Heat Capture:** Operate the LTDIs<sup>®</sup> process effectively using available waste heat under variable heating conditions (including surges/outages).
- 2) **Distillate Quality:** Produce consistent distillate below 500 mg/L of total dissolved solids.
- 3) **Operational Stability:** Demonstrate reliable, independent, and extended duration of water treatment operations without affecting the existing facilities at the compressor station.

Table A: Parameters for Phase 1 and Phase 2

Parameter	Phase 1	Phase 2
<b>Waste Heat Capture (MMBtu/hr)</b>	1.5 – 2.4	1.7 – 2.1
<b>Distillate Quality, average (Total Dissolved Solids, mg/L)</b>	388	303
<b>Operational Stability</b>	18.8 days of continuous operation, 16 hours per day	9.1 days of continuous operation, 16 hours per day

The 16-hour day measurement benchmark was established as an objective prior to the trial to allow for miscellaneous maintenance/modification activities as needed. Phase 1 included some downtime (eg. installing insulation on the heat source), but the majority of pilot days were 24h operating time, without the need for any regular maintenance.

Additional metrics used to evaluate the LTDIs<sup>®</sup> technology's performance were the electrical demand intensity, distillate production, thermal intensity, and thermal efficiency. Lower than expected heat input required high turndown, reducing the operating efficiency of the rotating equipment. The result was significantly higher than normal electrical demand intensity.

Heat input intensity was approximately 0.18 – 0.20 MMBtu/bbl. distillate produced. The intensity did not change significantly from fluctuating to steady heating conditions. Due to a damaged piece of equipment, we were forced to disconnect Stage 2 (evaporator/condenser pair), making that stage completely

inactive. A functioning Stage 2 was shown to reduce thermal intensity by approximately 0.03 MMBtu/bbl (15%) in previous tests.

Table B: Additional Parameters for Phase 1 and Phase 2

Parameter	Phase 1	Phase 2
<b>Distillate Production (bbl/d)</b>	225 (Nov 7-10) 250 (Nov 11-18)	238 (Dec 38)
<b>Thermal Input Required (MMBtu/bbldistillate)</b>	0.21 (Nov 7-10) 0.19 (Nov 11-18)	0.20 (Dec 3-8)
<b>Electrical demand intensity (kWh/bbldistillate)</b>	2.0 – 2.5	2.0 – 2.2

## 2 Project Objectives

The New Mexico Produced Water Research Consortium’s Request for Proposals sought bench-scale, pilot-scale, and field-scale research, development, and demonstration projects for the cost-effective treatment and reuse of produced water for different fit-for-purpose applications outside the oil and gas sector. The state of New Mexico is considering alternate pathways for produced water management and reuse that address: (1) Growing challenges of current produced water underground disposal practices that have been linked to environmental issues such as earthquakes and may not be sustainable in the long-term; and (2) Water scarcity and groundwater depletion, which are forecast to increase due to climate change.

CCR’s proposal addressed AOI-1; Research and demonstration of cost-effective and energy-efficient mobile or fixed treatment technologies compatible with both medium and high salinity produced water and their associated constituents, including:

- Treatment and desalination for constituent and contaminant removal
- Post-treatment for selective removal of constituents or water conditioning for fit-for-purpose use
- Minimization of concentrate, byproducts or liquid and solid wastes produced during the treatment processes
- Recovery of reusable water from produced water and its treatment residuals

### 2.1 Performance Objective Targets

The purpose of the pilot testing was to demonstrate the ability of the LTDis® technology to treat oilfield produced water with multi-stage distillation using available waste heat as an energy source. The distillate produced was required to be of acceptable quality for industrial or non-potable reuse. The water quality data collected during the Demonstration can be used by the NMPWRC to apply for a water discharge permit for similar produced water(s) at their locations.

Table C: Target Parameters for Phase 1 and Phase 2

Parameter	Phase 1 Target	Phase 2 Target
<b>Waste Heat Capture (MMBtu/hr)</b>	3.4	2.0
<b>Distillate Quality (Total Dissolved Solids, mg/L)</b>	< 250	< 500
<b>Operational Stability</b>	15 days of continuous operation, 16 hours per day	7 days of continuous operation, 16 hours per day

The targets for phases 1 and 2 were different because the pilot was designed around a heat availability of 3.4 MMBTU/hr and this level of heat was never available from the waste heat source. In Phase 1, we insulated the exhaust line to the Cain heat recovery unit (reducing heat loss). The heat provided to the CCR LTDIs unit fluctuated between 1.8 and 2.4 MMBTU/hr with an average of ~ 2.0 MMBTU/hr throughout Phase 1. No supplementary heat source was added in Phase 2 so the target heat was adjusted to 2.0 MMBTU/hr in Phase 2.

There were a few upset conditions in Phase 1 where there was carry over of high salinity brine from Evaporator 1 to Condenser 1 due to an inherent design flaw of the vapor channels in the D1 pilot unit identified through this pilot. The target was adjusted to be below 500 ppm for Phase 2 to adjust for the operating conditions that were encountered in Phase 1.

Additional operational parameters by which we evaluated the trial include:

- 1) Distillate Production, measured in barrels per day of distillate. This is dependent on the internal efficiency of the LTDIs® process and the external heat available.
- 2) Thermal efficiency of the operation, defined as the ratio of recovered heat to total heat.
- 3) Electrical Intensity of distillate production, defined as kilowatt-hours of electricity consumed per barrel of distillate produced.

## 2.2 Project Scope

The scope of work for the CCR team at the Demonstration included:

- Mobilization to the compressor location
- Setup and connection of the D1 unit to the waste heat system
- Cold water commissioning
- Waste heat and hot water commissioning/testing
- Full scale operations and testing in 12-hour and 24-hour shifts

The client's team was responsible for the supply and operation of the:

- H<sub>2</sub>S scavenger system

- Glycol system for heat input to heat exchanger 1 (HEX.1)
- Produced water feed to the D1 unit
- Effluent stream connections, including distillate and concentrated brine
- Utilities including fresh water, instrument air, and electrical power connection

H<sub>2</sub>S removal was only required for the D1 as a mitigation for safety risks associated with the vacuum line venting to the atmosphere. With a H<sub>2</sub>S venting mitigation plan (eg. relocation of the vent, or vapor destruction if required), a commercial plant could be designed to process H<sub>2</sub>S-containing water without significantly changing the construction materials.

Glycol requirements vary depending on the heat source configuration and proximity. Usually, the volume is <10 bbl. to fill the process piping but could exceed that for a large heat recovery system spanning multiple heat exchangers. The heating loop should include an atmospheric flash tank to allow for thermal expansion. The glycol/water mixture level can be varied depending on desired boiling point and heat capacity.

Utility fresh water is optional and is used to top-up the evaporative air cooler. Normally, we use produced distillate for the air re cooler, which consumes 20 – 50% of the produced distillate. In this case, no external fresh water is required.

Instrument air was supplied at approximately 85psi, with 3scf/min consumption.



Figure A. Compressor Location

\*Note: The pit looks shallow in Figure A, but that is water on top of the cover. Feedwater fed to site off 30,000 bwpd pipeline. Concentrate sent back downstream into pipeline to SWD.



Figure B. Layout of D1 Compressor Site

### 3 Results

The objectives were to use waste heat from compressor exhaust, produce high quality distillate of <500 ppm TDS, and demonstrate extended duration.

These objectives were demonstrated as achieved with the results:

- Successfully captured and utilized compressor waste heat while maintaining compressor operations
- Achieved turndown operations at ~2.0 MMBTU/hr low grade waste heat
- Delivered distillate treating 120,000-170,000 ppm feed:
  - Phase 1: avg. 338 mg/L TDS
  - Phase 2: avg. 303 mg/L TDS
- Consistently operated > 16hrs out of 24-hour day, averaging 70% uptime in Phase 1 and 88% uptime in Phase 2



Table D. Parameters with Target and Actual Results for Phase 1 and Phase 2

Parameter	Phase 1 Target	Phase 1 Actual	Phase 2 Target	Phase 2 Actual
<b>Waste Heat Capture (MMBtu/hr)</b>	3.4	1.5 – 2.4	2.0	1.7 – 2.1
<b>Distillate Quality (Total Dissolved Solids, mg/L)</b>	< 250	388	< 500	303
<b>Operational Stability</b>	15 days of continuous operation, 16 hours per day	18.8 days of continuous operation, 16 hours per day	7 days of continuous operation, 16 hours per day	9.1 days of continuous operation, 16 hours per day

Table E. Additional Parameters for Phase 1 and Phase 2

Parameter	Phase 1	Phase 2
<b>Distillate Production (bbl/d)</b>	225 (Nov 7-10) 250 (Nov 11-18)	238 (Dec 3-8)
<b>Thermal Input Required (MMBtu/bbl distillate)</b>	0.21 (Nov 7-10) 0.19 (Nov 11-18)	0.20 (Dec 3-8)
<b>Electrical demand intensity (kWh/bbl distillate)</b>	2.0 – 2.5	2.0 – 2.2

### 3.1 Produced Water and Blended Feed Water

CCR's LTDIs<sup>®</sup> Technology achieved desalination of a feed ranging from 120,000-170,000 ppm of produced water to below 500 ppm consistently over a 4-week operating period during Phase 1 and Phase 2. There was some co-distillation of dissolved organics into the distillate stream (5-10 ppm TOC) which included BTEX levels up to 0.5 ppm. A 2 stage, in-series activated carbon adsorption step was added in the final 2 days of the Phase 2 trial to demonstrate the capability of the system and prove that the technology can be applied with an activated carbon polishing step to achieve water quality that can be considered for surface discharge and potentially permitted under a NPDES permit. The activated carbon adsorption step reduced the BTEX levels to below 0.5 ppb. We are confident based on the learnings and the data collected from the pilot and the applications

for design into the commercial unit that the distillate quality will improve and a commercially available activated carbon polishing step for the water will meet the requirements for surface discharge quality.

As described in this section, Produced Water, pretreated by the client for H<sub>2</sub>S removal, was pumped to the Feed Tank where it was blended with Air Recooler Blowdown during Phase 1 operation.

Table F below compares the CWA analysis results of the:

- Raw, untreated Produced Water sampled before chlorine dioxide injection
- Phase 1 Feed Water sampled from the feed line to D1. Phase 1 feed was a blend of the chlorine dioxide-treated Produced Water and Air Recooler Blowdown
- Phase 2 Feed Water sampled from the feed line to D1. Phase 2 was only H<sub>2</sub>O<sub>2</sub>-treated Produced Water

For Phase 1, we used Chlorine dioxide as this equipment was already rigged in nearby for a different operation. The dosing equipment was dismantled following Phase 1 for deployment elsewhere. For Phase 2, Hydrogen Peroxide was used in Phase 2 because it was readily available for deployment.

Table F. Produced Water and Feed Quality

	Produced Water (1 sample)	Phase 1 Feed Water (Average 7 samples)	Phase 2 Feed Water (1 sample)
Sample Point	Before ClO <sub>2</sub> Injection	Feed to D1	Feed to D1
<b>Sodium (mg/L)</b>	40,610	29,378	38,163
<b>Calcium (mg/L)</b>	5,055	3,875	4,555
<b>Sulfate (mg/L)</b>	150	546	160
<b>Chloride (mg/L)</b>	77,300	52,643	86,200
<b>TDS (mg/L)</b>	126,649	95,803 (87,700 to 192,442)	132,253

Unexpectedly, Air Recooler Blowdown and Produced Water in the Feed Tank significantly increased the sulfate concentration due to freshwater source variability, but also decreased the TDS of the Phase 1 Feed Water by dilution. Phase 2 Feed Water did not have any added Air Recooler Blowdown water and was similar to the Produced Water as the only difference is the H<sub>2</sub>O<sub>2</sub> injection.

As a result of commingled blowdown (not produced) water, there was some sulfate scale accumulation on the evaporator inlet screen. This was only noted when the high sulfate (>1,000 mg/L) fresh water (not produced water) was commingled with the produced water feed. Scale accumulation was not significant enough to affect the evaporator function. Once this scale was cleaned and the high-sulfate fresh water was purged, no further scaling was observed in the evaporators through the entire duration of Phase 2. No issues with oxidized iron were observed. Because evaporation is taking place off water droplets (not from heating on a metallic surface), LTDIs® is much less susceptible to scaling than conventional thermal technologies. However, scaling is still possible if incompatible brines are commingled and fed into the system. Scaling and corrosion were not noted on critical rotating equipment and instruments/sensors.

Fresh Water quality is further discussed in section 5.8.

## 4 Review of Materials and Methods Used

### 4.1 Technology Overview

The D1 unit is a modular temporary water treatment system consisting of a base container package with an integrated control cabinet/operator panel, a tower package, a distillate tank skid, a pumping skid, an evaporative air re cooler skid, a field office building, and a storage container for other accessories and spares. The D1 unit was engineered and fabricated overseas for non-sour service.

The customer installed all external/support facilities, including the exhaust heat capture system, sour pre- treatment chemical injection, electricity connections, and all water influent/effluent piping connections to the D1. After treatment with D1, the client received the distillate water and all the concentrated waste streams back to their operation.

The modular plant design uses non-corroding materials and durable standardized components. Primary components of the LTDIs® plants include pressure vessels and evaporator spraying systems. These components are manufactured from fiber reinforced plastics (FRP) and/or stainless steel. Piping within the plant is made of FRP, polypropylene, or stainless steel. The external plate heat exchangers are made of titanium or stainless steel.

In contrast to many other thermal desalination technologies, LTDIs® is tolerant of highly variable feed water composition without the need for extensive pre- or post-treatment typically required for scale control.

LTDIs® plants can accommodate variations in the plant load, ranging from 25% to 100% of plant design capacity. The process is self-adjusting and adapts continuously to changing temperature conditions and can balance fluctuations in the heat load. The amount of water produced is always proportional to the amount of heat provided. As LTDIs® is a thermal process, a heating and cooling source are required to maintain a thermal differential for steady plant operation.

The LTDIs® technology is an innovative low temperature evaporation technology operating under vacuum conditions. The D1 unit has been designed for temporary installation and data collection necessary to optimize a commercial scale plant. Over 155 sensors are included to



measure flow, temperature, pressure, level, and conductivity over the four process loops. These process loops include the:

- 1) Heating/glycol loop
- 2) Evaporator loop
- 3) Condenser loop
- 4) Cooling loop

The LTDIs® process operates within automatically controlled temperature and pressure conditions for each of the evaporator and condenser vessels to facilitate efficient vaporization and condensation. The vacuum system maintains thermodynamic non-equilibrium within the plant which enables the driving force for continuous distillate production.

#### 4.2 Tie-In to the Client's Facility

The D1 was rigged into the produced water delivery system at an existing deep well injection site. Produced water was transported via High Density Polyethylene (HDPE) pipelines to a tank on the compressor site (Pilot location). This five-hundred-barrel (500 bbl) storage tank on site was filled regularly, providing a residence time of at least 12 hours for the produced water feed. Upon treatment of the distillate, the concentrated brine streams were recombined at offsite frac tanks and transported via HDPE lines to the Saltwater Disposal (SWD) site.

The LTDIs® system feedwater was pre-treated with an H<sub>2</sub>S scavenger in the form of Chlorine Dioxide (ClO<sub>2</sub>) and Hydrogen peroxide (H<sub>2</sub>O<sub>2</sub>) respectively for Phases 1 and 2 by the client. The only other pre-treatment of the water upstream of the D1 was a basket strainer with a 1/16" mesh screen.

The external heating source was a glycol circulation media that recovered waste exhaust heat from a compressor station. The supply temperature of the glycol was normally 80 - 90°C (175 – 195°F).



Figure C. D1 at the Pilot Location with Feed Tank (black tank) and Fresh Water Tank (blue tank)

The section below focuses on the key streams used in normal operation to describe the interfaces between the client facility and the LTDIs<sup>®</sup> pilot.

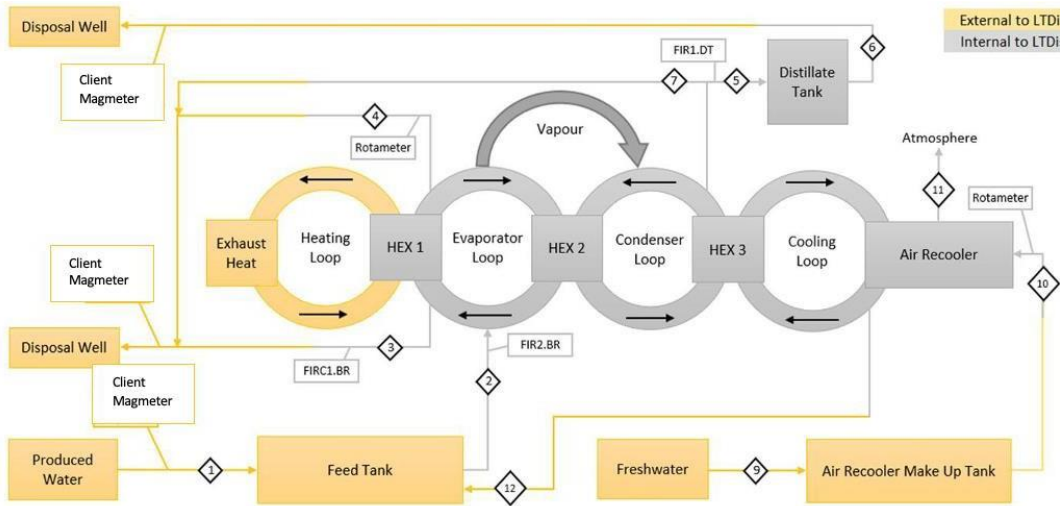


Figure D. Client Tie-In Simplified Block Flow Diagram - Phase 1

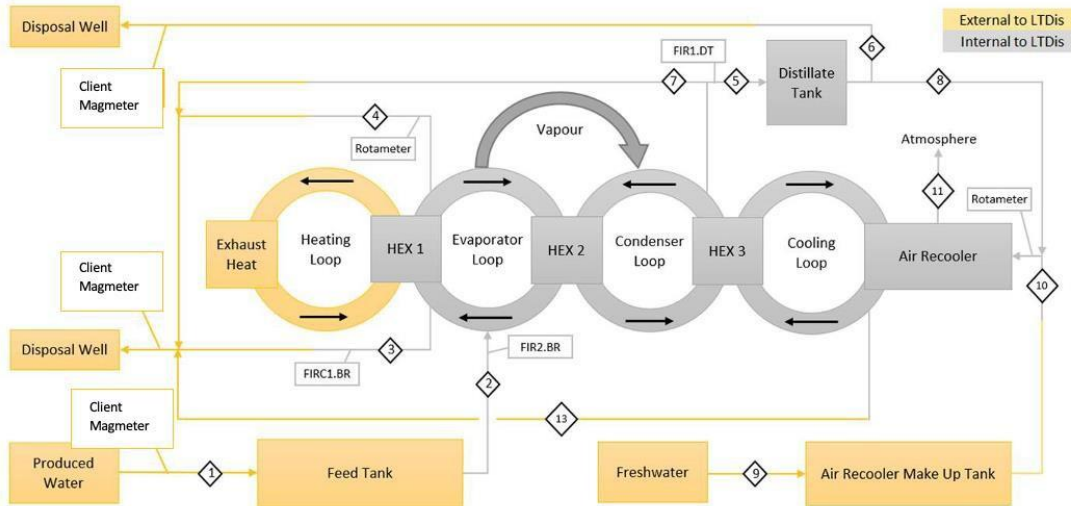


Figure E. Client Tie-In Simplified Block Flow Diagram – Phase 2

Table G. Normal Operation Stream Description

Stream	Name	From	To	Internal Flow Measurement
1	Produced Water	Client ClO <sub>2</sub> Injection	Feed Tank	no
2	Feed Water	Feed Tank	Evaporator Loop	In-Line Flowmeter
3	Brine Concentrate	Evaporator Loop	Disposal	In-Line Flowmeter
4	Misc. Brine	Siphon Discharge	Disposal	Rotameter
5	Distillate	Condenser Loop	Distillate Tank	In-Line Flowmeter
6	Distillate Discharge	Distillate Tank	Disposal	No
7	Off Spec Distillate	Condenser Loop	Disposal	No
8	Distillate for Air Recooler Make Up	Distillate Tank	Air Recooler	Rotameter
9	Fresh Water	Client Truck-In	Air Recooler Make Up Tank	Truck Manifest
10	Fresh Water for Air Recooler Make Up	Air Recooler Make Up Tank	Air Recooler Reservoir	Rotameter
11	Air Recooler Evaporative Losses	Air Recooler	Atmosphere	no

#### 4.2.1 Stream 1: Produced Water

A slip stream of Produced water was pumped by client operations. The produced water flow was treated with chlorine dioxide (ClO<sub>2</sub>) for H<sub>2</sub>S removal in Phase 1 and hydrogen peroxide for Phase 2 before being pumped to the pilot site. The slip stream sent to the LTD<sub>is</sub><sup>®</sup> pilot was diverted from a point downstream of the chlorine dioxide treatment, but upstream of the bulk scale inhibitor treatment to a feed tank. The level of the feed tank was monitored, and the tank was periodically refilled as required. Produced water was first circulated through D1 on November 3<sup>rd</sup>, 2021.

For Phase 1, we used Chlorine dioxide as this equipment was already rigged in nearby for a different operation. The dosing equipment was dismantled following Phase 1 for deployment elsewhere. For Phase 2, Hydrogen peroxide was used in Phase 2 because it was most readily available for deployment.

#### 4.2.2 Stream 2: Feed Water

Feed Water in the feed tank was introduced into the evaporator loop to replace the volume that was evaporated, as determined by the level in the standpipe upstream of PL.E7. The flow rate and volume totalizer of this stream was measured by in-line flowmeter FIRC.DT.

In Phase 1, Feed Water was a mixture of Produced Water (Stream 1) and Air Recooler Blowdown (Stream 12). In Phase 2, Feed Water consisted of Produced Water (Stream 1) only. A food grade silicone antifoam was dosed manually into the Feed Tank occasionally when foaming was observed in the evaporator vessels. The only pre-treatment of the Feed Water was a basket strainer with 1/16" perforation on the feed line.

The quantity of food grade silicone antifoam added was very small, 1-5 ppm in solution as needed, and no significant carryover of the antifoam into the distillate was observed.

#### 4.2.3 Stream 3: Brine Concentrate

Brine Concentrate was discharged to the disposal well from the evaporator loop based on a high conductivity setpoint. The flow rate and volume totalizer of this stream was measured by in-line flowmeter FIRC1.BR. Brine Concentrate analysis is discussed below in section 6.1. No filtration was completed following Brine Concentrate discharge, but this is an important consideration for future installations if any ions approach solubility. After the water cools upon leaving the plant, it is possible that precipitates could form in the waste stream.

Miscellaneous Brine consisted of brine and potentially solids in the siphons. It was pumped by Brine Discharge Pump PL1.BR from the siphons to the disposal well based on a timer. The volume totalizer of this stream was measured by a rotameter, which was recorded manually. On Nov. 15<sup>th</sup>, 2021, this rotameter was removed due to a connection failure, therefore no flow data on this stream after that time is available.

No significant precipitation in the brine discharge streams was observed.

#### 4.2.4 Stream 5: Distillate

Distillate accumulated in the condenser loop was removed through control valve VC.H3 by Condensate Transfer Pump PL1.DT, based on the level measured in Condenser 2. The distillate was sent to the Distillate Tank if the conductivity is below the off-spec distillate conductivity setpoint. The flow rate and volume totalizer of this stream entering the Distillate Tank was measured by in-line flowmeter FIR1.DT. Distillate analysis is discussed below in 5.6.

#### 4.2.5 Stream 6: Distillate Discharge

The distillate in the Distillate Tank was pumped to the disposal well when the level in the Distillate Tank was high. There was no onboard flow measurement on this stream.

#### 4.2.6 Stream 7: Off Spec Distillate

Off Spec Distillate consisted of distillate that was diverted from the Distillate Tank due to high conductivity. The diversion occurred upstream of the Distillate Flow Meter FIR1.DT, therefore, this stream was not measured or included in the distillate production data.

#### 4.2.7 Stream 8: Distillate for Air Recooler Make Up

The Air Recooler is an evaporative cooler that uses the air-water heat exchanger to remove excess heat from the condensate loop by ambient air evaporation driven by an axial fan.

The distillate from the Distillate Tank was used for Air Recooler make up towards the end of Phase 2 (starting from Dec. 6<sup>th</sup>, 2021, 19:45). This was done to test the feasibility of stripping ammonia from the distillate using the Air Recooler. The volume totalizer of Air Recooler Make Up was measured by a rotameter and data was recorded manually.

#### 4.2.8 Stream 9: Utility Fresh Water (Optional)

Fresh Water was trucked in to replenish the Air Recooler Make Up Tank. The volume was recorded on the truck manifests. Fresh Water analysis is discussed below in section 5.8.

Note that the LTDIs<sup>®</sup> process can use distillate as the cooling loop fluid. However, to minimize the safety and environmental risks of evaporating the unverified distillate in the open atmosphere, Fresh Water was used instead for most of the pilot.

#### 4.2.9 Stream 10: Fresh Water for Air Recooler Make Up (Optional)

Fresh Water was transferred by gravity flow from the Air Recooler Make Up Tank to the Air Cooler Reservoir when the level in the reservoir was low. The volume totalizer of Air Recooler Make Up was measured by a rotameter and data was recorded manually. Make Up Water Demand is discussed in section 6.7.

#### 4.2.10 Stream 11: Air Recooler Evaporative Losses

A portion of water in the cooling loop evaporated as it flowed through the Air Recooler. The volume of cooling water lost to evaporation was not measured but an estimate is provided in section 6.7.

#### 4.2.11 Stream 12 and Stream 13: Air Recooler Blowdown

Evaporative losses in the Air Recooler increased the TDS concentration of the cooling loop water. To prevent scaling and corrosion, Air Recooler Blowdown occurred based on a conductivity setpoint. Air Recooler Blowdown flow and volume totalizer were not measured.

In Phase 1, Air Recooler Blowdown water was sent to the Feed Tank (Stream 12). High sulfate content in the blowdown water led to an acceleration of scaling in Heat Exchanger 1 (HEX.1). The external HDPE piping was modified by the client before the start of Phase 2 to allow for Air Recooler Blowdown to be sent to disposal instead (Stream 13).

### 4.3 Post Treatment

On December 8<sup>th</sup>, 2021, an activated carbon filter system was installed on the distillate discharge using two, forty-two-gallon drums connected in series as can be seen in Figure F. Table H. shows distillate water quality before and after polishing with a carbon filter. The carbon filter system was able to achieve a high degree of VOC and TPH removal.





Figure F. Activated Carbon System

Table H. Distillate Water Quality Before and After Carbon Filter

Parameters (mg/L)	Distillate Pre- Carbon Filter Representative Sample	Distillate Post Carbon Filter 12/08/21
<b>Benzene</b>	0.501	<0.000214
<b>Toluene</b>	0.548	<0.000500
<b>Ethylbenzene</b>	0.0214	<0.000515
<b>Xylenes</b>	0.377	<0.000330
<b>TPH (C6 to C12)</b>	1.51	<0.840
<b>TPH (&gt;C12 to C28)</b>	1.53	<0.819
<b>TPH (&gt;C28 to C35)</b>	<0.860	<0.819
<b>TPH (C6 to C35)</b>	3.04	<0.840
<b>Fluorene</b>	<0.00163	<0.00163
<b>Naphthalene</b>	0.00993	<0.000542
<b>Phenanthrene</b>	0.00145	<0.00142
<b>2-Nitrophenol</b>	0.007	<0.00167
<b>Phenol</b>	0.026	0.000693

No testing of air emissions was performed during testing as it was outside the scope of the pilot. For a commercial design, air emissions will be managed to meet local regulations.

CCR did not conduct any analysis for Ammonia when the samples were sent to a 3<sup>rd</sup> party laboratory. The client conducted its own independent sampling and analysis of the feed and the distillate in Phase 1, and it was noted that the feed contained ammonia ranging from 450 – 600 ppm and the distillate contained ammonia ranging from 45-60 ppm. The distillate samples provided to both NM state and the Aris Water Solutions towards the end of Phase 2 indicated ammonia levels ranging from 35-50 ppm.

## 5 Discussion of Results and their Significance

### 5.1 Water Quality

Throughout the operation, samples were collected from the Feed, Evaporator Loop, and Distillate streams. These samples were sent to a local 3<sup>rd</sup> party Laboratory analysis. These tests included:

- Complete Water Analysis (CWA)
- Total Petroleum Hydrocarbon (TPH)
- Volatile Organic Carbon Species (VOCs)

The total number of samples collected for each analysis is summarized in the table below.

Table I. Phase 1 Quantity of Samples Collected for Analysis

Phase 1			
Analysis	Evaporator Loop	Feed	Distillate
<b>CWA</b>	16	8	21
<b>TPH</b>	7	3	9
<b>VOCs</b>	1	2	8

Table J. Phase 2 Quantity of Samples for Analysis

Phase 2			
Analysis	Evaporator Loop	Feed	Distillate
<b>CWA</b>	1	1	4
<b>TPH</b>	0	0	0
<b>VOCs</b>	1	1	1

The significance of the results demonstrates the possibility of treating high saline produced water with the use of waste heat as the sole energy source.

During Phase 2, four distillate, one evaporator loop, and one feed sample were taken by Oilfield Labs of America (OLA) for analysis. Two distillate samples (pre and post carbon) were collected

and sent to APEX Laboratory for detailed environmental analysis. Independent samples of distillate were also provided to NMPWRC and Aris Water Solutions. The detailed analytical data from those samples are not currently available.

## 5.2 Time and Downtime during 24-hour Operations

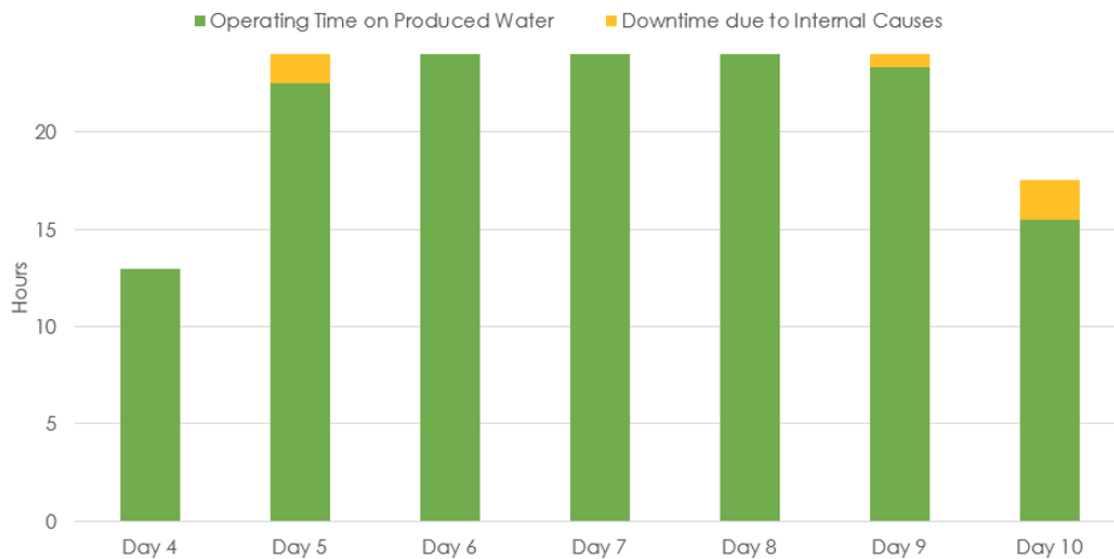


Figure G. Operating Time and Downtime during 24-hour Operations - Phase 2

The cause for downtime in Phase 2 operation included troubleshooting of salinity spikes (day 5), failure and repair of an air compressor (Day 9) and flushing of the condenser loop due to brine crossover (Day 10). Days 1 to 3 of Phase 2 comprised of pre-cleaning and flushing of residues left at the end of Phase 1 testing. Several weeks elapsed between the two phases.

Carryover occurred when heat upsets caused a slight rise in the brine level in Evaporator 1 vessel. In order to fit the plant tower into a shipping container, the vapor channel nozzle was low in the vessel. A simple upward piping extension on future plants will solve this issue.

## 5.3 Stable Flow Rates

Evaporator loop flow rate, condenser loop flow rate, and distillate production trends during Phase 2 operation are shown in Figure H, demonstrating continuous stable operation from Day 5 to Day 9.



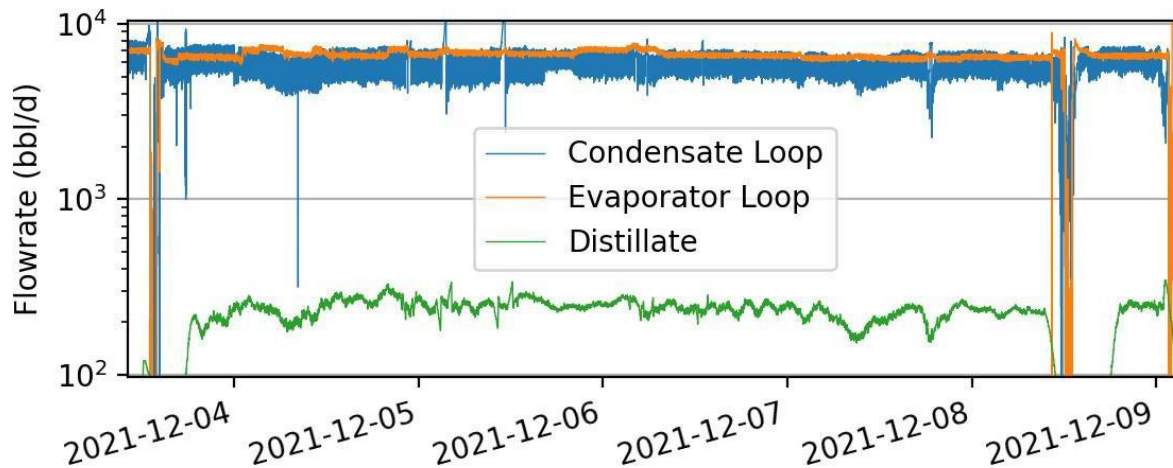


Figure H. Phase 2 Operation - Stable Flow Rate

#### 5.4 Definition of Total Dissolved Solids

Depending on the intended use of the analysis, Total Dissolved Solids (TDS) are analyzed by different methods. The methods are not equal and do not produce comparable results. The different methods of TDS analysis are measuring different contaminants:

##### 1) TDS Gravimetric Method 2540C, EPA160.1

The gravimetric method is TDS analysis in a bulk solid detection method. The sample is filtered with a 2  $\mu\text{m}$  glass fiber filter followed by evaporation at 180°C (356°F). The residue is weighed with a precision analytical balance. This method reports mass only, there is no characterization of the TDS components.

##### 2) TDS Summation 1030E

TDS by summation method 1030E (Factored Summation) is the sum of cations and anions and includes a correction factor of 0.6 applied to bicarbonate anions to account for decomposition of this anion when heated. This method is used by environmental laboratories to characterize the components of TDS.

##### 3) TDS Summation – No Standardized Method (Complete Summation)

Petroleum Laboratories will report TDS as the total summation of all dissolved contaminants including dissolved gases and all cations and anions with no correction factors applied. The complete summation method is problematic because it over-calculated the true solids present in the water post evaporation making the results unapplicable to environmental work.

Within the scope of this trial, TDS was analyzed by 2 different labs:

- APEX Labs - Gravimetric TDS Analysis EPA 160.1
- OLA - TDS Summation (Complete summation of all dissolved contaminants)

The TDS results reported by the two labs are not comparable as they use different methods. As seen in the analysis, bicarbonate anions ( $\text{HCO}_3^-$ ) are the dominant anion in the distillate. When the correction factor specified by method 1030E is not applied, the calculated TDS significantly

overestimates the TDS of the water. Bicarbonate anions will decompose, when heated, to become carbonate anions, carbon dioxide gas, and water.

For consistency, TDS values referenced in this report were re-calculated from lab-reported cation, anion, and bicarbonate values using TDS Calculation Method 1030E for Factored Summation.

Plants currently under development will include instruments to measure density directly, and field measurements will include use of a mud balance.

### 5.5 Produced Water and Blended Feedwater

Produced Water, pretreated by the client for H<sub>2</sub>S removal, was pumped to the Feed Tank where it was blended with Air Recooler Blowdown during Phase 1 operation.

Table K below compares the CWA analysis results of the:

- Raw, untreated Produced Water sampled before chlorine dioxide (ClO<sub>2</sub>) injection
- Phase 1 Feed Water sampled from the feed line to D1. Phase 1 feed is a blend of the chlorine dioxide-treated Produced Water and Air Recooler Blowdown
- Phase 2 Feed Water sampled from the feed line to D1. Phase 2 is only H<sub>2</sub>O<sub>2</sub>-treated Produced Water

Table K. Produced Water and Feed Water Quality

	Produced Water (1 sample)	Phase 1 Feed Water (Average 7 samples)	Phase 2 Feed Water (1 sample)
Sample Point	Before ClO <sub>2</sub> Injection	Feed to D1	Feed to D1
<b>Sodium (mg/L)</b>	40,610	29,378	38,163
<b>Calcium (mg/L)</b>	5,055	3,875	4,555
<b>Sulfate (mg/L)</b>	150	546	160
<b>Chloride (mg/L)</b>	77,300	52,643	86,200
<b>TDS (mg/L)</b>	126,649	95,803 (87,700 to 192,442)	132,253

As expected, the blending of Air Recooler Blowdown and Produced Water in the Feed Tank significantly increased the sulfate concentration but also decreased the TDS of the Phase 1 Feed Water. Phase 2 Feed Water was similar to the Produced Water as the only difference is the H<sub>2</sub>O<sub>2</sub> injection. Fresh Water quality is further discussed in section 5.8.

CCR team is confident that the brine concentrate can be increased to 240,000-260,000 ppm

due to the presence of high sodium and calcium salts. Individual scenarios should be analyzed, however, to determine which precipitates will form first. The LTDis® plant can handle precipitated particles, but this may be problematic for equipment downstream of LTDis®, like disposal wells.

## 5.6 Distillate

Distillate produced by the plant was monitored throughout the pilot. There were six 24-hour steady state operating days in Phase 1 and four 24-hour steady state operating days in Phase 2 where consistent distillate quality was achieved. The distillate water quality was stable (Conductivity < 500  $\mu\text{S}/\text{cm}$ ) except during upset conditions where Brine Concentrate was carried over into the condenser loop. The upset events and spikes in conductivity can be seen in Figure I and Figure J. Several D1 software updates were made after Phase 1, reducing the likelihood of carryover from Evaporator 1 to Condenser.

Carryover occurred when heat upsets caused a slight rise in the brine level in evaporator 1 vessel. To fit the plant tower into a shipping container, the vapor channel nozzle was low in the vessel. A simple upward piping extension on future plants will solve this issue. The D1 and plants under development have real-time redundant instruments to prevent the discharge of off-spec distillate.

## 5.7 Conductivity and TDS

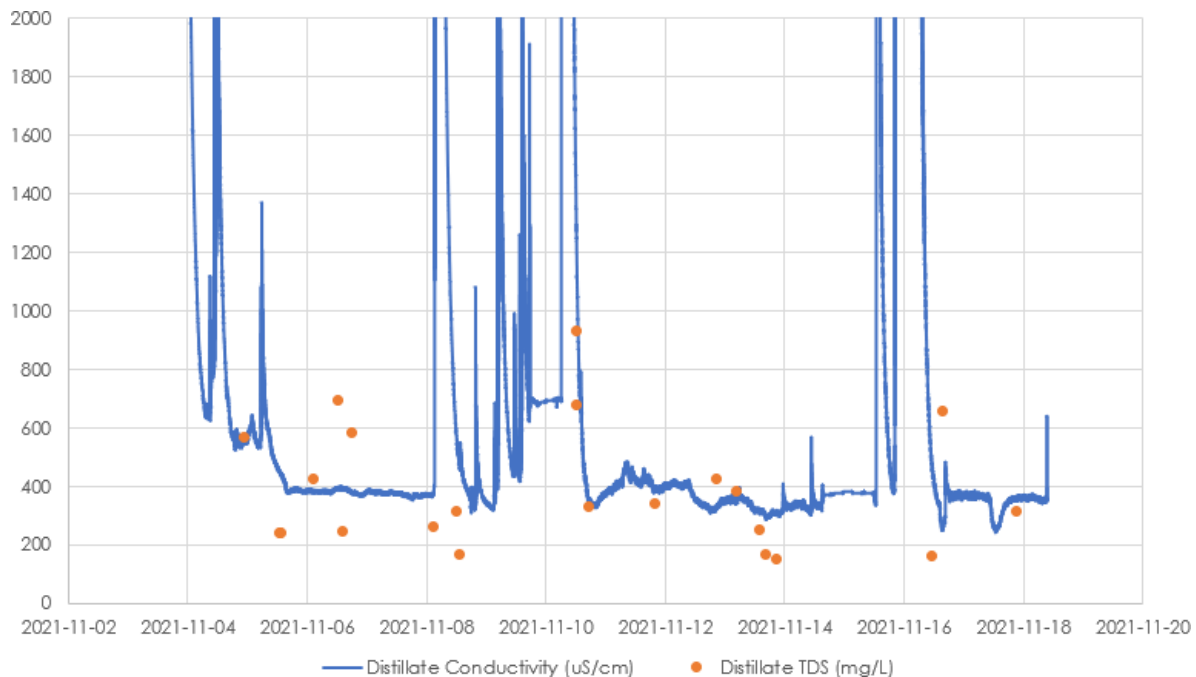


Figure I. Phase 1 Distillate Conductivity and TDS

It is important to note that conductivity from in-line sensors was not necessarily measured on the same water as the TDS samples sent to external labs. TDS measurements were

sometimes higher, sometimes lower than would be predicted by conductivity measurements. It is possible that different ionic and non-ionic constituents at different times caused different relationships between conductivity and TDS.

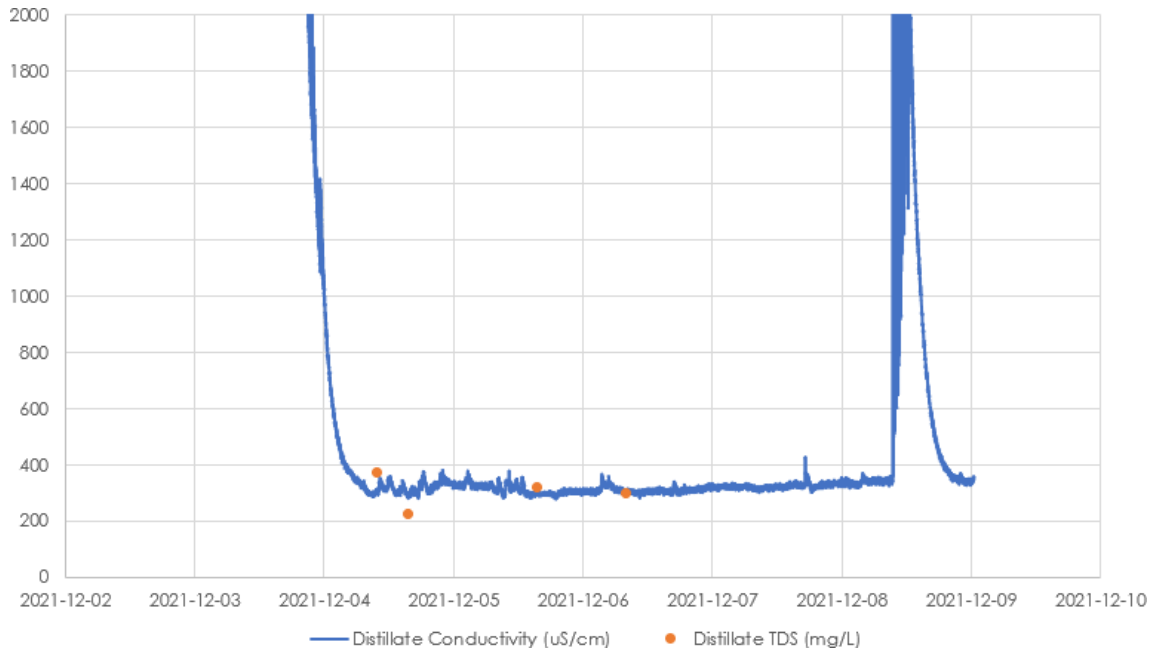


Figure J. Phase 2 Distillate Conductivity and TDS

Table L. Distillate Total Dissolved Solids

	Average/ Std Dev (mg/L)	Min (mg/L)	Max (mg/L)
<b>Phase 1 and 2</b>	376 ± 190	152	929
<b>Phase 1</b>	388 ± 202	152	929
<b>Phase 2</b>	303 ± 54	223	372
<b>Steady Operating Days</b>	349 ± 146	152	694

TPH results from 3rd party lab analysis of the distillate compared to the feed and evaporator loop are shown in Table M.

Table M. Feed, Evap Loop, and Distillate TPH

	Total Petroleum Hydrocarbon Average ± Std Dev (mg/L)
<b>Feed</b>	73.50 ± 78.94
<b>Evaporator Loop</b>	12.17 ± 5.64
<b>Distillate</b>	11.33 ± 3.30

Average Benzene, Toluene, Ethylbenzene, and Xylene (BTEX) data from the 3rd party lab VOCs analysis of the distillate compared to the feed and evaporator loop is shown in Figure K. Most of the VOCs in the feed is carried over in the distillate.

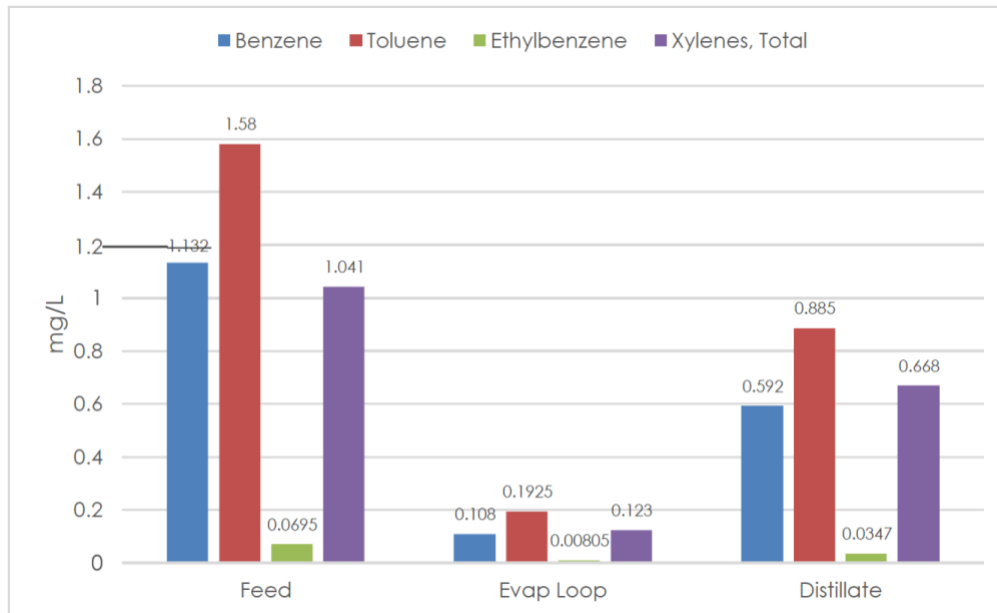


Figure K. Feed, Evap Loop, and Distillate BTEX

Ammonia results from 3rd party lab analysis in the distillate and the feed are shown in Table N.

Table N. Feed and Distillate Ammonia

	Ammonia as N (mg/L) - Dec 7, 2021
<b>Feed</b>	500
<b>Distillate</b>	40

The source of the ammonia content is not known at the current time, it is possible it came from biocide use. Biocides particularly such as Dazomet, Dibromo-nitrilopropionamide, N-Bromosuccinimide, Brionopol etc. can be sources of compounds that can create ammonia. The client conducted its own independent sampling and analysis of the feed and the distillate in Phase 1, and it was noted that the feed contained ammonia ranging from 450 –600 ppm and the distillate contained ammonia ranging from 45-60 ppm.

## 5.8 Freshwater

Utility Fresh Water at the beginning of Phase 1, was supplied from an unspecified source and was trucked in for the Air Cooler Make Up. It was switched to potable city water halfway through Phase 1 when lab tests showed high sulfate content. Table O summarizes the water

quality of the two Fresh Water samples from the Fresh Water Tank. Both sources show high sulfate concentrations. Although potable city water was expected to have much less sulfate and TDS, it was found to be similar to the unspecified Fresh Water.

Table O. Fresh Water Analysis

	Unspecified Fresh Water	Potable City Water	Potable City Water
<b>Sample Point</b>	Fresh Water Tank	Fresh Water Tank	Fresh Water Tank
<b>Sample Date</b>	October 30, 2021	November 10, 2021	November 17, 2021
<b>Sodium (mg/L)</b>	653	650	569
<b>Calcium (mg/L)</b>	791	703	468
<b>Sulfate (mg/L)</b>	1,020	1,600	1,020
<b>Chloride (mg/L)</b>	1,200	2,210	1,440
<b>TDS (mg/L)</b>	4,728	5,478	3,774

Unexpected variations in the quality of the fresh water source resulted in 4,000 – 5,000 ppm of dissolved solids, including incompatible ions which caused some scale deposition. For this reason, it is preferable to use internally recycled distillate for the cooling loop. In this way, we do not have to rely on the quality of external sources of fresh water for evaporative cooling and can minimize the risk of scaling-associated downtime.

## 5.9 Thermal Demand

### 5.9.1 Input and Output Flows (Feed Water and Brine Discharge)

The average feed rate was 606 bbl./d in Phase 1 and 633 bbl./d in Phase 2. The average brine discharge flow rate was 333 bbl./d in Phase 1 and 370 bbl./d in Phase 2. These are flow rates as measured at the inlet/outlet of D1 during the operating time and extrapolated to 24-hour.

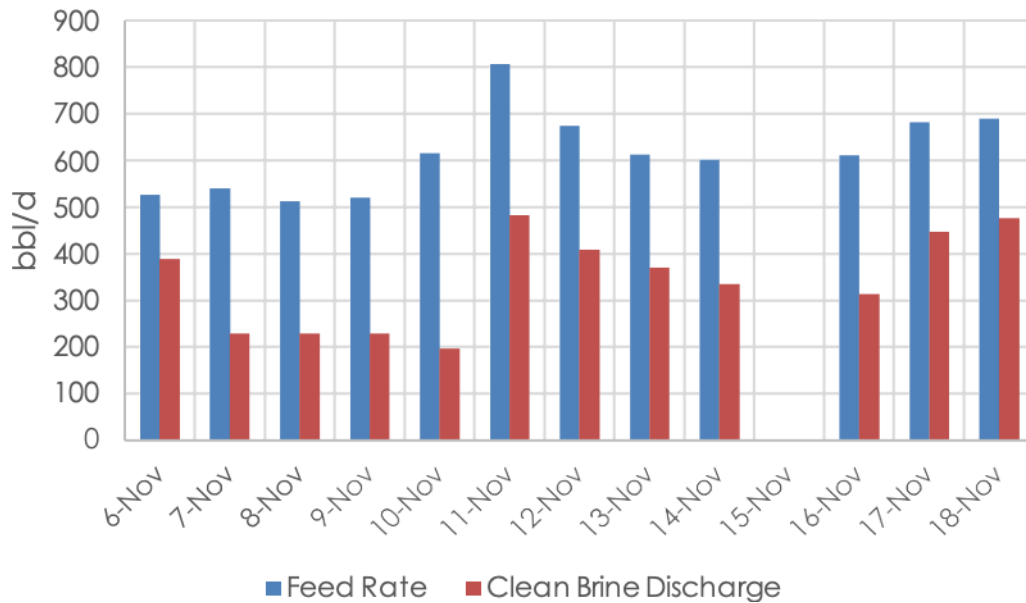


Figure L. Phase 1 Feed and Brine Discharge Rate

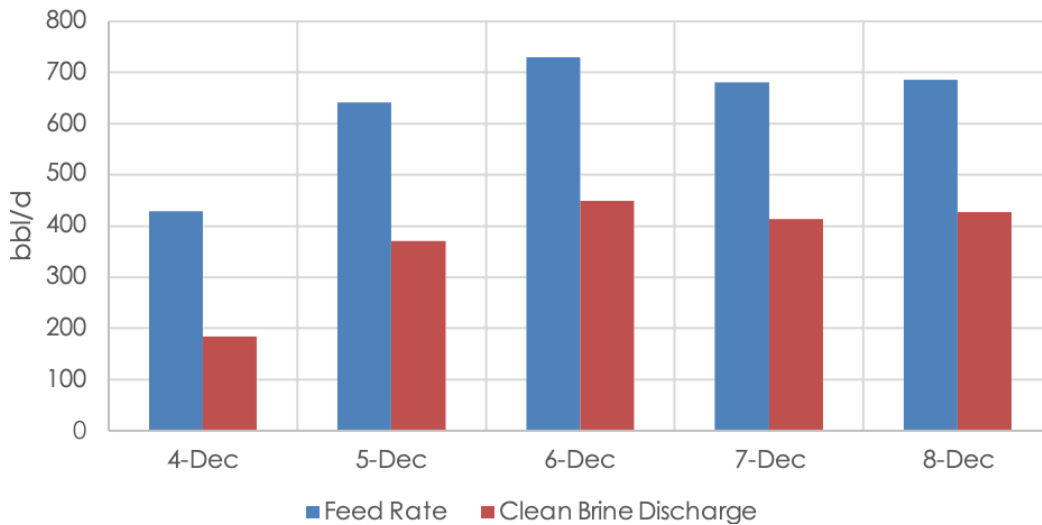


Figure M. Phase 2 Feed and Brine Discharge Rate

Note that “Clean Brine” in this case is the concentrated brine discharged, not including the small amount of “Miscellaneous Brine” discharged from the evaporator siphons.

### 5.9.2 Distillate Production

Distillate production accounts for water generated by the D1 unit that also meets conductivity targets of < 500  $\mu\text{S}/\text{cm}$ . If the distillate did not meet a conductivity of < 500  $\mu\text{S}/\text{cm}$ , then the distillate was not metered, and it went to disposal. The average distillate production was 238 bbl/d in Phase 1 and 244 bbl/d in Phase 2. These are production rates during operating time

extrapolated to 24-hour.

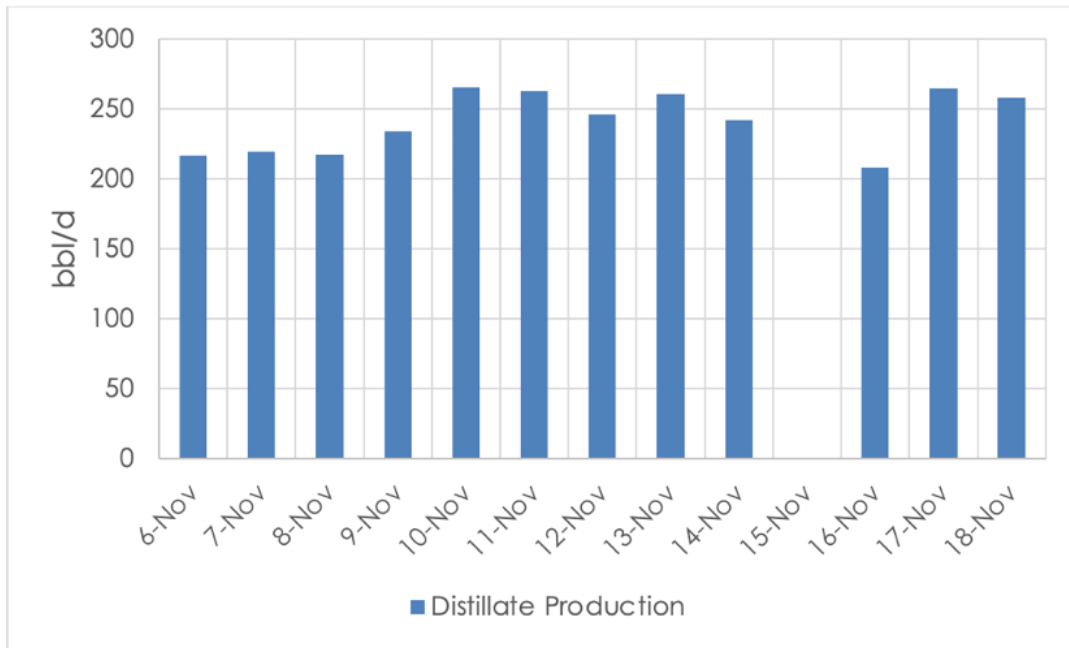


Figure N. Phase 1 Distillate Production

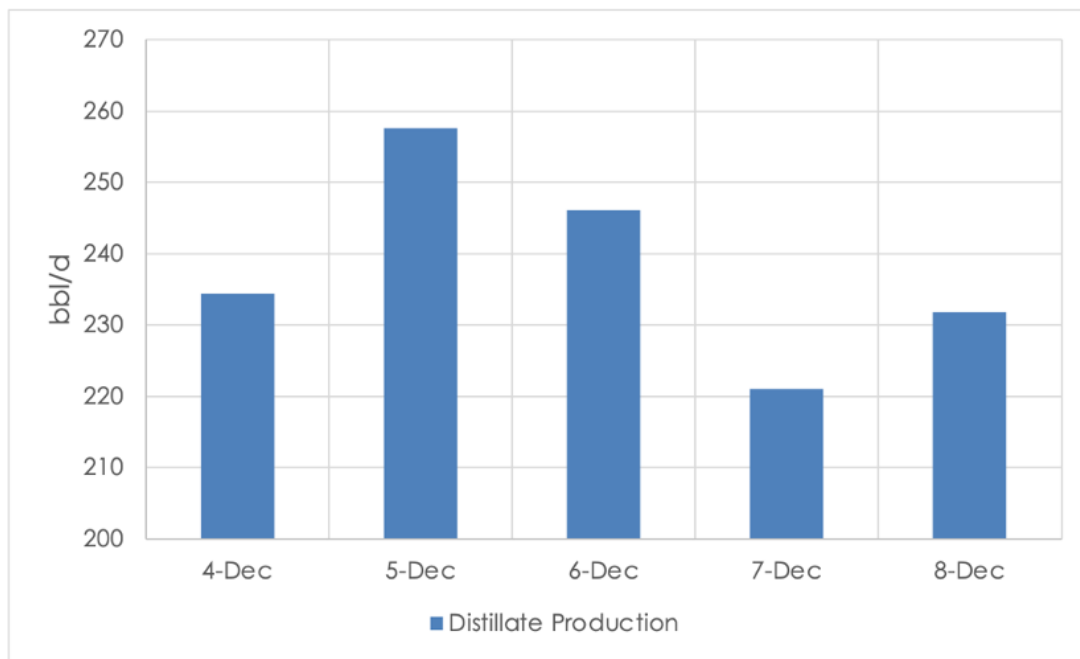


Figure O. Phase 2 Distillate Production

### 5.9.3 Heat Input

The main external source of energy for the D1 plant during the demonstration was waste heat captured from the gas compressor through a Cain exhaust heat recovery system. The original



plan was for 3.4 MMBtu/hr to be provided to the D1. The actual heat supplied ranged from 1.5-2.4 MMBtu/hr and 1.7-2.1 MMBtu/hr for Phases 1 and 2, respectively. We attributed the lower-than-expected heat available to either a lower compressor load than expected (due to field production conditions) or lower than expected recovery efficiency of the exhaust heat recovery system.

While we observed variable heating conditions, the heat intensity (heat input required per barrel of distillate produced) was quite constant. The process continued to function with a distillate production directly proportional to the heat input. The average thermal intensity for Phases 1 and 2 was 0.18 – 0.20 MMBtu/hr.

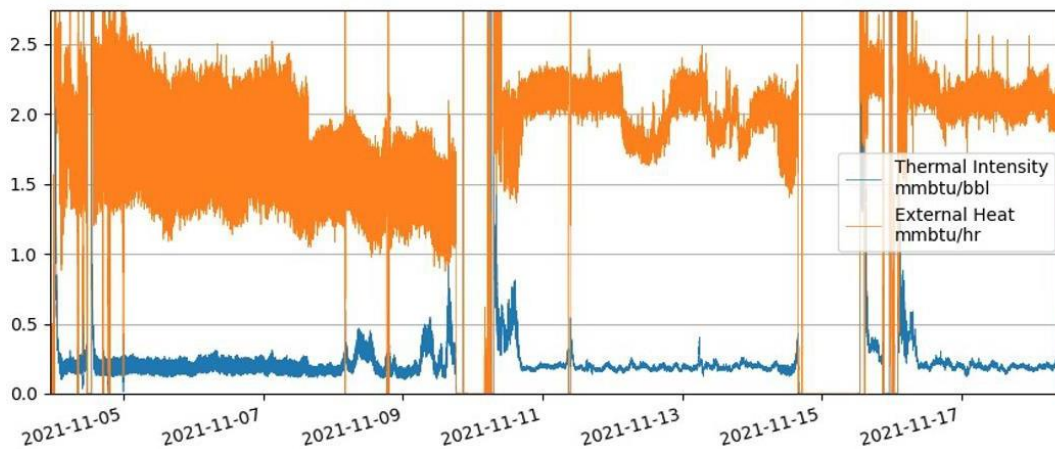


Figure P. Phase 1 Thermal Input and Thermal Intensity of D1 Operation

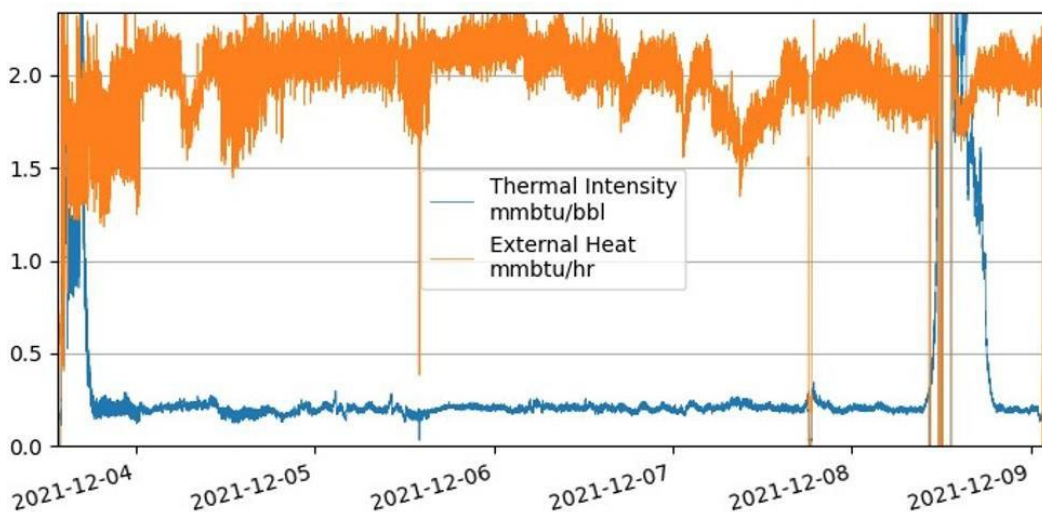


Figure Q. Phase 2 Thermal Input and Thermal Intensity of D1 Operation

Note that in the data shown for both phases, the thermal heat input has been corrected. From temperature and flowrate instruments, it was calculated that the heat-transfer rate was ~20% higher on the glycol side vs. the brine side of HEX.1. To investigate this discrepancy, data from non-operational times was examined, and it was found that TIRC1.H1 (the hot glycol inlet temperature probe) was reading an ambient temperature of approximately 2°C (4°F) warmer than all the other probes in the D1 plant. Therefore 2°C (4°F) was subtracted from each data point to calculate the external heat delivered to the plant. There is still a 10% discrepancy between the two temperature values. Since the heat was balanced between the evaporator and condenser loops based on measured values, it was determined that the external heat provided in HEX.1 may be overestimated by up to 10%.

### 5.9.4 Operating Expenses

Operating Expenses for LTDIs<sup>®</sup> are mostly comprised of utilities; external heat to drive the evaporation process and electricity for rotating equipment and instruments. Let's consider a scenario where an operator uses natural gas to heat, at a cost of \$6.00 per MMBtu., and electricity from the grid at \$0.07/kWh. Utility OPEX in this case would be ~\$0.98 per bbl. of distillate, and ~\$0.49 per bbl. feed. Waste heat significantly improves OPEX; zero-OPEX waste heat in the previous scenario would result in an overall OPEX of \$0.07 per bbl. of distillate and \$0.04 per bbl. of feed. For produced waters containing high levels of ammonia or volatile organic compounds, minor polishing (eg. activated carbon, air sparging, or oxidation) to meet specific discharge requirements. A range of \$0.05-\$0.10 cents per bbl for water polishing costs for commercial units may be required in such cases.

Assumptions for the above OPEX scenarios are LTDIs<sup>®</sup> consumption of 0.15MMBtu and 1.1kW-h per bbl. distillate, and 2:1 concentration of brine (eg. 140kppm feed to 280kppm concentrate). CCR is currently working on a commercial next generation design that will include the learnings from this demonstration.

**Heating Cost (US\$ per MMBtu)**

	\$/BBL	\$ -	\$ 0.50	\$ 1.00	\$ 1.50	\$ 2.00	\$ 2.50	\$ 3.00	\$ 3.50	\$ 4.00	\$ 4.50	\$ 5.00	\$ 5.50	\$ 6.00
<b>Electricity Cost (US\$ per kW-h)</b>	\$ -	\$ -	\$ 0.08	\$ 0.15	\$ 0.23	\$ 0.30	\$ 0.38	\$ 0.45	\$ 0.53	\$ 0.60	\$ 0.68	\$ 0.75	\$ 0.83	\$ 0.90
	\$ 0.02	\$ 0.02	\$ 0.10	\$ 0.17	\$ 0.25	\$ 0.32	\$ 0.40	\$ 0.47	\$ 0.55	\$ 0.62	\$ 0.70	\$ 0.77	\$ 0.85	\$ 0.92
	\$ 0.04	\$ 0.04	\$ 0.12	\$ 0.19	\$ 0.27	\$ 0.34	\$ 0.42	\$ 0.49	\$ 0.57	\$ 0.64	\$ 0.72	\$ 0.79	\$ 0.87	\$ 0.94
	\$ 0.06	\$ 0.06	\$ 0.14	\$ 0.21	\$ 0.29	\$ 0.36	\$ 0.44	\$ 0.51	\$ 0.59	\$ 0.66	\$ 0.74	\$ 0.81	\$ 0.89	\$ 0.96
	\$ 0.08	\$ 0.09	\$ 0.16	\$ 0.24	\$ 0.31	\$ 0.39	\$ 0.46	\$ 0.54	\$ 0.61	\$ 0.69	\$ 0.76	\$ 0.84	\$ 0.91	\$ 0.99
	\$ 0.10	\$ 0.11	\$ 0.18	\$ 0.26	\$ 0.33	\$ 0.41	\$ 0.48	\$ 0.56	\$ 0.63	\$ 0.71	\$ 0.78	\$ 0.86	\$ 0.93	\$ 1.01
	\$ 0.12	\$ 0.13	\$ 0.20	\$ 0.28	\$ 0.35	\$ 0.43	\$ 0.50	\$ 0.58	\$ 0.65	\$ 0.73	\$ 0.80	\$ 0.88	\$ 0.95	\$ 1.03
	\$ 0.14	\$ 0.15	\$ 0.22	\$ 0.30	\$ 0.37	\$ 0.45	\$ 0.52	\$ 0.60	\$ 0.67	\$ 0.75	\$ 0.82	\$ 0.90	\$ 0.97	\$ 1.05
	\$ 0.16	\$ 0.17	\$ 0.25	\$ 0.32	\$ 0.40	\$ 0.47	\$ 0.55	\$ 0.62	\$ 0.70	\$ 0.77	\$ 0.85	\$ 0.92	\$ 1.00	\$ 1.07
	\$ 0.18	\$ 0.19	\$ 0.27	\$ 0.34	\$ 0.42	\$ 0.49	\$ 0.57	\$ 0.64	\$ 0.72	\$ 0.79	\$ 0.87	\$ 0.94	\$ 1.02	\$ 1.09
	\$ 0.20	\$ 0.21	\$ 0.29	\$ 0.36	\$ 0.44	\$ 0.51	\$ 0.59	\$ 0.66	\$ 0.74	\$ 0.81	\$ 0.89	\$ 0.96	\$ 1.04	\$ 1.11

**OPEX Matrix (\$US per barrel distillate)**

Figure R. Heating Cost (US\$ per MMBtu)

## 6 Principal Findings and Conclusions

### 6.1 Concentrated Brine Quality

Brine Concentrate within the system was monitored throughout the pilot test. The evaporator loop recirculated brine through the heat exchangers and evaporators until the total dissolved solids concentration setpoint (measured as conductivity) was reached, at which point Brine Concentrate was discharged back to the client. Periodically, samples of the brine were taken to measure TDS so that conductivity and TDS could be correlated.

The average rate of brine recirculation rate was 47 m<sup>3</sup>/h during the pilot. The TDS concentration in the Brine Concentrate was typically kept at < 200,000 mg/L to prevent precipitation of solids, although the TDS reached as high as 282,320 mg/L during Phase 2. In Phase 1, the average TDS in the evaporator loop was 127,392 mg/L ± 67,577. Figure R shows the TDS of the feed, evaporator loop, and distillate. Note that in Phase 2, the distillate TDS remained in the same range as Phase 1 even though evaporator loop TDS was doubled.

At full-scale deployment, we plan to concentrate up to 250 – 280k ppm (depending on solubility of specific ionic constituents), which will be enabled by the use of direct in-line measurement of density. Our toroidal conductivity sensors were not reliable at such levels of salinity, so we kept a buffer to be conservative in preventing unexpected precipitation.

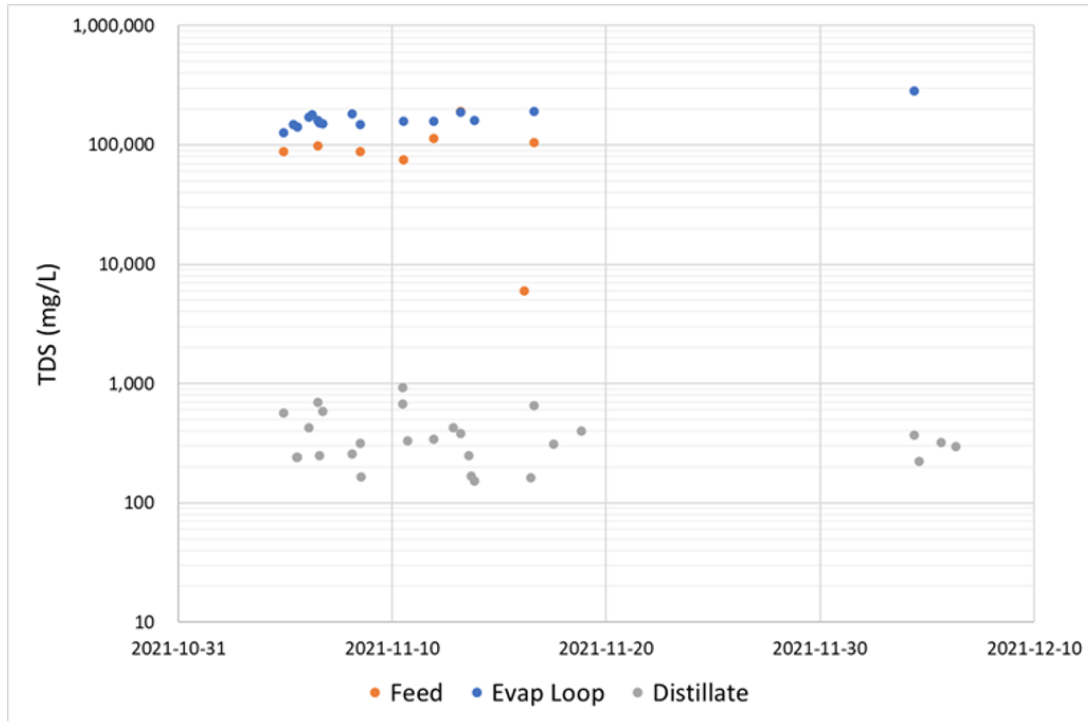


Figure S. Total Dissolved Solids of Feed, Evap Loop, and Distillate

## 6.2 Misc. Brine

The siphons between the evaporators provided a low velocity point for any suspended solids to settle and be discharged from the evaporator recirculation loop. The siphons were opened several times during testing and no buildup of solids was observed.

## 6.3 Distillate Water Quality

Once steady state was achieved, D1 was successfully able to treat high salinity feed water streams during Phases 1 and 2. The produced water feed of up to 170,000 ppm TDS was

encountered particularly in Phase 2. The salinity of the distillate stream ranged from 220-700 ppm throughout the steady state conditions excluding upsets, with a repeatedly achieved lower limit of 152 ppm TDS. The objectives of desalinating a feed water stream ranging in salinity > 120,000 ppm down to < 500 mg/L TDS was achieved successfully.

There was carryover of dissolved Volatile Organic Compounds (VOCs) in the distillate stream in both Phases 1 and 2. The 2-stage activated carbon polishing step was able to reduce the dissolved VOCs and in particular Benzene, Toluene, Ethylbenzene, Xylene (BTEX) at least 3 orders of magnitude and below 0.5 parts per billion (ppb). The inclusion of the activated carbon step had a significant impact on heavy metal reductions as well. Based on the trial results, activated carbon is an excellent option for post-distillation polishing for VOC removal. CCR is also testing electrochemical oxidation methods for future direction of post-treatments.

The distillate water quality is expected to be better in the next generation units with lower salinity and even lower dissolved organics concentration in the unpolished distillate. We plan to accomplish higher-quality distillate by incorporating simple piping geometry changes based on lessons learned from the D1 plant. This should lead to better more efficient process overall and lower make up volumes and lower volumes of blowdown.

## 6.4 Waste Heat and Thermal Intensity

The LTDis<sup>®</sup> demonstration unit successfully demonstrated the capability to utilize low grade waste heat (< 99°C or < 210°F) from engine exhaust for desalination of very high saline produced water, up to 170,000 ppm, to an average TDS of less than 500mg/L in the distillate. The heat available to the demonstration plant averaged 2 MMBtu/hr representing a turndown of approximately 70% of nameplate capacity. Process stability was maintained during such high turndown and there were zero negative impacts on the process or its equipment from sudden unexpected heat outages. Additionally, there was no observable difference in distillate quality during even large variations in heat input. The main findings concerning heat input were found to be:

- 1) Consistent thermal intensity of ~0.2 MMBtu/bbl of distillate across a wide range of heating conditions (1.5–2.4 MMBtu/hr, including several unexpected outages).
- 2) No observed change in distillate quality due to variance (or unexpected outage) of the heat source.
- 3) Large operable turndown range (1.5–4.0 MMBtu/hr) with little impact on thermal efficiency,

but turndown did affect electrical intensity due to poorer efficiency of rotating equipment during high turndown.

The thermal intensity of 0.2 MMBtu/bbl. is analogous to a gain output ratio (GOR) of ~2, a measure commonly referred to for Multiple-Effect Distillation (MED) units. A carefully designed

MED plant can reach GOR of >6. Our GOR is extrapolated from the performance of the current prototype, which we believe can be significantly improved. Our goal is to provide future plants with energy efficiencies closer that of MEDs/MSFs, with existing differentiating features of smaller capex, footprint, scaling risk, and larger operational versatility.

There are significant learnings from the pilot around the limitations of the vacuum system as designed in the D1. These limitations resulted in an operating pressure that was ~250mbar in the last (coldest) evaporator and had a significant impact on thermal performance. The vacuum pump design in future units will allow for an operating pressure of 100-150 mbar which will reduce non-condensable gas content and therefore improve overall efficiency. Operating at a lower bottom-end temperature is expected to increase primary heat recovery and decrease thermal production intensity.

The pilot unit was originally fabricated for approximately 1,500-2,000 bwpd of feed. Subsequent modifications and piping changes in 2020 led to a capacity of approximately 1200 bwpd. Due to heat availability of 2.0 MMBTU/hr vs. the design nameplate of 4.8 MMBTU/hr for the pilot the feed capacity was further reduced to approx 650-700 bwpd. The system was, however, still able to demonstrate the process capabilities at a significant turndown.

## 6.5 Thermal Efficiency

Defined as the ratio of total heat (glycol heat plus heat recovered) to glycol heat, we observed thermal efficiencies in the range of 2 (1.6–2.4), depending on inclusion or exclusion of the feed-warming effect. This was constant under variable heat input (1.5 – 2.4 MMBtu/hr).

Both measures indicate that the pilot LTD<sup>is</sup>® unit was recovering approximately as much heat from internal condensation recovery as externally from glycol/exhaust waste heat. Significant opportunities, such as better leak-tightness and simplified piping layouts, exist to improve both thermal efficiency and intensity.

$$Q_{Evaporation} = Q_{HEX1} + Q_{HEX2}$$

$$Q_{HEX1} = \text{External Heat Source}$$

$$Q_{HEX2} = \text{Recovered Energy}$$

$$TEF = \frac{Q_{Evaporation}}{Q_{HEX1}}$$

Figure T. Calculation for Thermal Efficiency

There are two methods for calculating the recovered energy, or  $Q_{HEX2}$ . Because the feed water was cooler than the brine loop, HEX.2 will remove slightly more heat due to a higher temperature difference between the cooler brine side (measured between HEX.2 outlet temperature and TIR3.H2A) than the hotter condensate side (measured between HEX.2 outlet temperature and TIRC.E7).

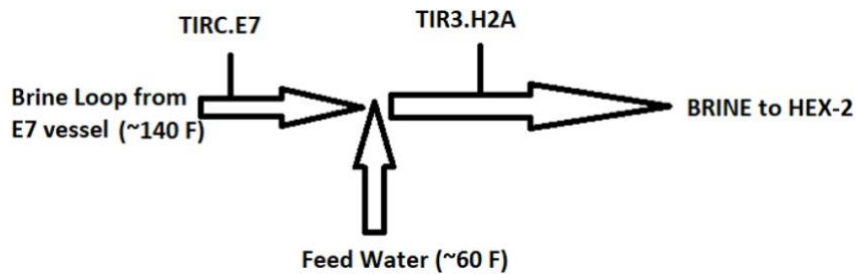


Figure U. Cool Feed Water Entering the Warm Brine Loop

Since there is no industry-standard practice for reporting the heat recovery or the thermal efficiency, thermal efficiency using both methods of calculating recovered energy is presented (see Figure S and Figure T).

In the future, larger vapor channels and/or increased stage counts will reduce the brine loop temperature. Bringing the bottom of the brine loop temperature down will increase heat transferred from the condensate side, and the thermal efficiency will rise. Similarly, the difference between the two calculation methods will reduce as the brine loop temperature becomes closer to the feed temperature.

During Phase 1, efficiencies of approximately 2.0 (including the feed-warming effect), and 2.5 (not including the feed-warming effect) were observed.

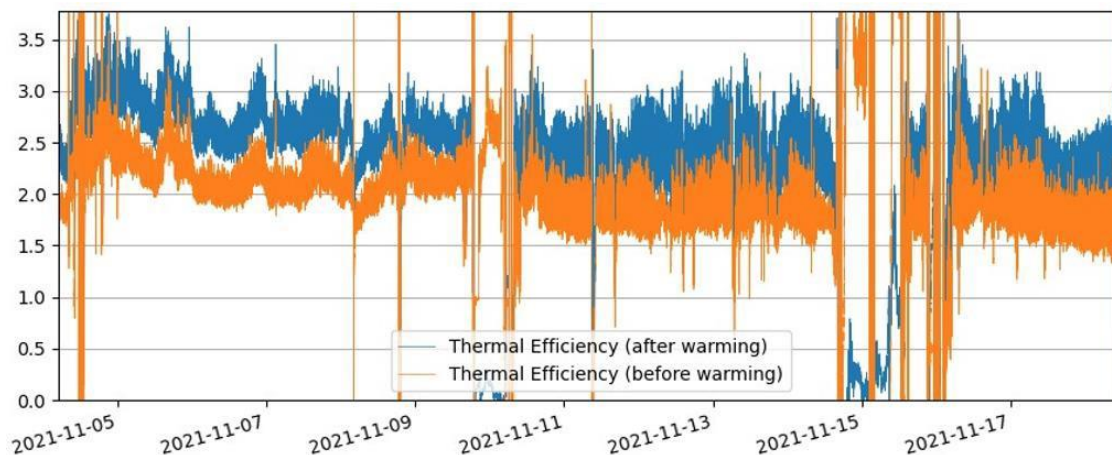


Figure V. Thermal Efficiency Phase 1

A slow downward trend of thermal efficiency during Phase 1 was observed. The exact reason(s) for this are not known for certain, but it's possibly due to a combination of more saline feedwater, changing heat conditions, or gradual scaling of the HEX.2 plates, reducing its heat-transfer capacity.



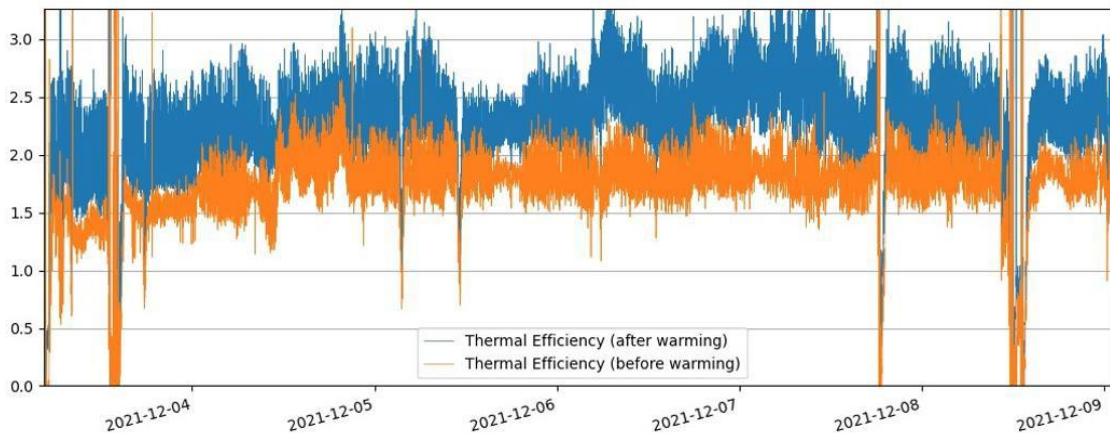


Figure W. Thermal Efficiency Phase 2

Phase 2 thermal efficiencies were observed to be 1.6 and 2.4, on average, as measured by including and excluding the feed-warming effect, respectively. Though there was not an opportunity to inspect HEX-2 (it was captured by tightly packed piping), we noticed that the thermal efficiency in Phase 2 did not reduce over the test as observed in Phase 1. It is possible this can be explained by less scaling (which would reduce heat exchanger plate efficiency) as a result of not commingling incompatible brines. Only produced water was treated in Phase 2.

## 6.6 Electrical Intensity

During Phase 1 and 2 trials at the customer's site, the electrical intensity of 2.2 kWh/bbl of distillate was observed. This was mostly because the pumps were operating at a significant turndown, and therefore a lower efficiency (the main evaporator recirculation pump, for example, operated at approximately 40% of rated load).

The circulation of the brine loop and the condensate loop represents the single largest electrical requirement for producing distillate. The brine loop has one pump, and the condensate loop has three pumps which all move water from the bottom of the tower to the top to gravity flow through the respective evaporator or condenser vessels. As displayed in the figure below, there was a steep decrease in pumping efficiency as conditions moved left of the design flowrate.

During the pilot, the average brine circulation rate was 47 m<sup>3</sup>/hr, resulting in a calculated efficiency of ~30%. Additionally, theoretical pump work calculations were performed and compared with the actual electrical draw through the VFDs, showing an efficiency of 33%. During a previous testing phase, the electrical intensity was observed to be ~1.1 kWh/bbl of distillate, and other opportunities exist to further reduce the electrical intensity.

Vapor Press:

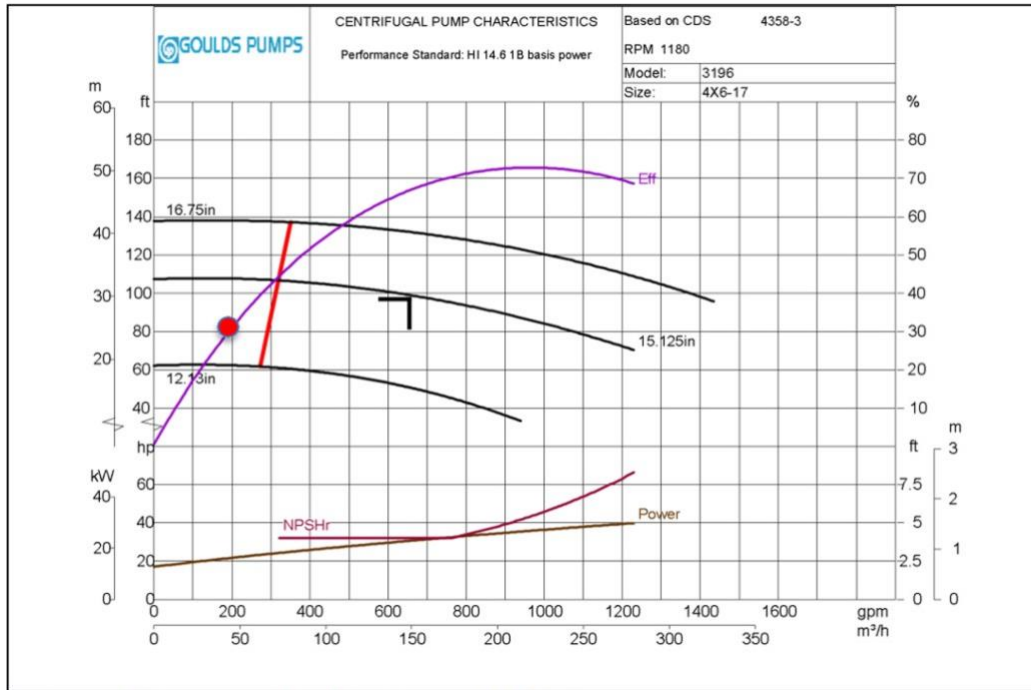
**Notes:** 1. Curve shown is at ambient temperature conditions.


Figure X. Pump Curve for PL.E7 - Brine Recirculation Pump

Table P summarizes the actual thermal energy requirements (MMBtu/bbl) and the electrical requirements (Kwh/bbl) distillate production for Phase 1 and 2 and the estimated treatment costs are summarized below based on \$0.07/Kwh basis. We have not assumed a cost for thermal energy as one hundred percent waste heat was used in Phases 1 and 2.

Table P. Summary of Thermal &amp; Electrical Requirements

	Phase 1 Nov 7 - 10	Phase 1 Nov 11 - 18	Phase 2 Dec 3 - 8
<b>Distillate Production (bbl./hr)</b>	9.4	10.4	9.9
<b>Thermal Req. (MMBtu/bbl.)</b>	0.21	0.19	0.20
<b>Electrical Req. (kwh/bbl.)</b>	2.13	2.48	2.19
<b>Treatment Cost (\$/bbl.)</b>	\$0.15*	\$0.17*	\$0.15*

\*Based on \$0.07/Kwh basis power costs



## 6.7 Utility Water Demand

The volume of Air Recooler make up is summarized in Table Q. Note that this utility water is optional, and distillate can be internally recycled and used in the evaporative cooler.

Table Q. Volume of Air Recooler Make Up

	Phase 1	Phase 2	Total
<b>Air Recooler Make Up (bbl)</b>	2635	882	3517
<b>Distillate Production (bbl)</b>	2918	1277	4195
<b>Air Recooler Make up per Distillate Production</b>	0.90	0.69	0.84

Note that the above make up water demand is not optimized due to the high TDS in the “Fresh Water” provided, requiring frequent blowdown to prevent corrosion and precipitation in the Air Recooler system.

Evaporative loss in the air recooler was estimated to be 71.4 bbl/day. This was done by stopping the blowdown sequence for an hour and calculating the Air Recooler Make Up volume as measured by a rotameter.

The next generation plants are expected to use the distillate for the Air Recooler make up requirements. We are also working on additional adaptations for air recooler requirements depending on the client needs (i.e. some clients may want a higher final distillate recovery for beneficial use etc vs some clients wanting a minimal distillate recovery and maximum evaporation).

## 7 Recommendations for Additional Research or Applications

Future research and development will focus on improvements to thermal efficiency and electrical efficiency to allow future iterations of this technology to produce more water from less energy input. Though utilization of waste heat was proven through this pilot, future pilots will focus on utilization of renewable thermal inputs such as solar thermal and geothermal. This technology also has applications outside of Oil and Gas, such as in municipalities, solution mining, and industrial sectors.

## 8 Summary and Lessons Learned

The Low Temperature Distillation (LTDis®) technology was used successfully to demonstrate the ability to treat oilfield produced water at the compressor facility (the Demonstration). For the Demonstration, the D1 Pilot LTDis® Unit (D1) was driven by low grade exhaust waste heat from a Caterpillar 3608 compressor.

This technology can be a significant driver for lowering carbon emissions in oil and gas

operations and can produce a significantly lower cost source of “freshwater production” in areas with water scarcity.

The pilot tests successfully demonstrated the D1’s ability to treat oilfield produced water with a Total Dissolved Solids (TDS) ranging from 100,000-170,000 ppm, for at least fifteen 24-hour operating days, while producing a distillate stream of less than 500 ppm TDS. This was achieved using only waste heat from a single gas compressor. The heat availability for the Demonstration ranged from between 1.5 and 2.4 MMBtu/hr, which was lower than the planned minimum of 3.4 MMBtu/hr, resulting in an overall lower rate of produced water treated and distillate generated. Plant performance proved to be consistent throughout the range of water salinities received during the Demonstration, including a feed salinity spike up to 192,000 mg/L.

Future commercial LTDIs<sup>®</sup> plants ranging from 2,500 bbl/day to 10,000 bbl/day are being engineered to have improvement in thermal and electrical efficiencies of at least 30% based on learnings and analysis from the Demonstration.

The Demonstration results and lessons learned from Phases 1 and 2 have provided significant and high-quality data to support the successful design and deployment of a full-scale application of the LTDIs<sup>®</sup> Technology at multiple water treatment locations.

CCR recognizes the need for validation of full-scale design improvements. Some improvement initiatives are obvious enough just from data and options and observations, while we are planning to validate others through lab/shop testing.

### 8.1 Thermal Intensity and Efficiency

During Phase 1 and 2 of the Demonstration, the thermal efficiency was 1.6 – 2.4. This is determined by several key factors in the operation and design of the plant including the following:

- Temperature difference across heat exchangers
- Completeness of evacuation of non-condensable gasses
- Vertical temperature differential (Temperature at the top of Evaporator 1, minus the temperature at the bottom of Evaporator 7)

The vertical temperature differential is of key importance. The vertical temperature differential represents the mass and energy transfer from the evaporator loop to the condenser loop in the form of distillate production. A higher vertical temperature differential increases the temperature difference between cold brine and hot condensate. Such a condition promotes greater heat recovery from condensate through HEX.2, and thus higher thermal efficiency. An average vertical temperature differential of 21°C (70°F) was observed. One key reason for this was that stage 2 was disconnected due to a previously damaged piece of equipment. Eliminating non-performing stages would improve thermal performance to ~0.15 MMBtu/bbl., as the D1 plant has done in previous tests prior to the equipment damage.

To achieve greater vertical differential and thermal performance beyond 0.15 MMBtu/bbl., the following measures can be taken:

- Improve leak-tight sealing of piping, allowing for better vacuum-sealing performance
- Optimize vacuum pump selection to draw initial pressure lower and reduce non-condensable gas content (requires simultaneous improvement in leak-tightness). This will allow for a reduced bottom-end operating temperature and a higher vertical temperature spread
- Alternate materials specifications to accommodate higher Stage 1 inlet temperatures (if available/expected from waste heat)

## 8.2 Electrical Efficiency

Operating at turndown (~1.8 vs 4.0 MMBtu/hr nominally, approximately 60% turndown), while operationally stable, had a significant impact on electrical intensity (>2 kWh/bbl distillate). The electrical intensity observed during the pre-deployment shop trials was ~50% lower than that observed during the Demonstration (1.1 kWh/bbl distillate).

Further improvements in electrical intensity could be made, by:

- Right-sizing of pumps to operate in high-efficiency zone under expected conditions
- Right-sizing of VFDs to reduce relative waveform-conversion power loss
- Reduce vacuum pump load by improving piping leak-tightness and enlarging vapor channels to lower friction loss

## 8.3 Scaling

During the pilot, scaling was observed in Heat Exchanger 1 as a result of commingling groundwater with high sulfate levels with the produced water, and extremely low circulating rates through our plate and frame heat exchangers while operating at 3:1 turndown. While LTDis® technology avoids scaling problems associated with evaporating on metallic surfaces, care should be taken in future plants to avoid commingling incompatible brines.

For likely scaling locations in the plate and frame heat exchangers, next-generation plants will be designed based on available heat to optimize fluid velocity across metallic surfaces. For turnaround situations, simple clean-in-place connections will be included to simplify the cleaning of critical elements, like heat exchangers, if required.

## 8.4 Conductivity Spikes

While we consistently produced distillate with 300 – 400mg/L total dissolved solids, some upset conditions resulted in levels closer to 1,000mg/L. These, however, occurred during small spillover events and were subsequently cleaned out of the system, and high-quality distillate was restored within a few hours.

To fit in a standard shipping container, some of the stages in the pilot unit have horizontal and upward vapor channels. As a result, brine can crossover from the evaporators to condensers through horizontal and downflow vapor channels if the level in the evaporator is too high due to

a pressure or temperature upset. This cross over resulted in conductivity spikes during operation.

Commercial system design will incorporate minor piping geometry modifications to minimize crossover risk.

### 8.5 Pre- and Post-Treatment

A feed tank at the plant inlet will be required as part of several LTDis® process requirements. Once inlet conditions to the plant have been determined as part of pre-feed engineering, an updated tank

design will be developed to account for floatation and settling of expected contaminants and any design upset conditions.

After outlet requirements from the plant have been developed during pre-feed engineering, post treatment of outlet streams will be completed to ensure downstream compatibility.

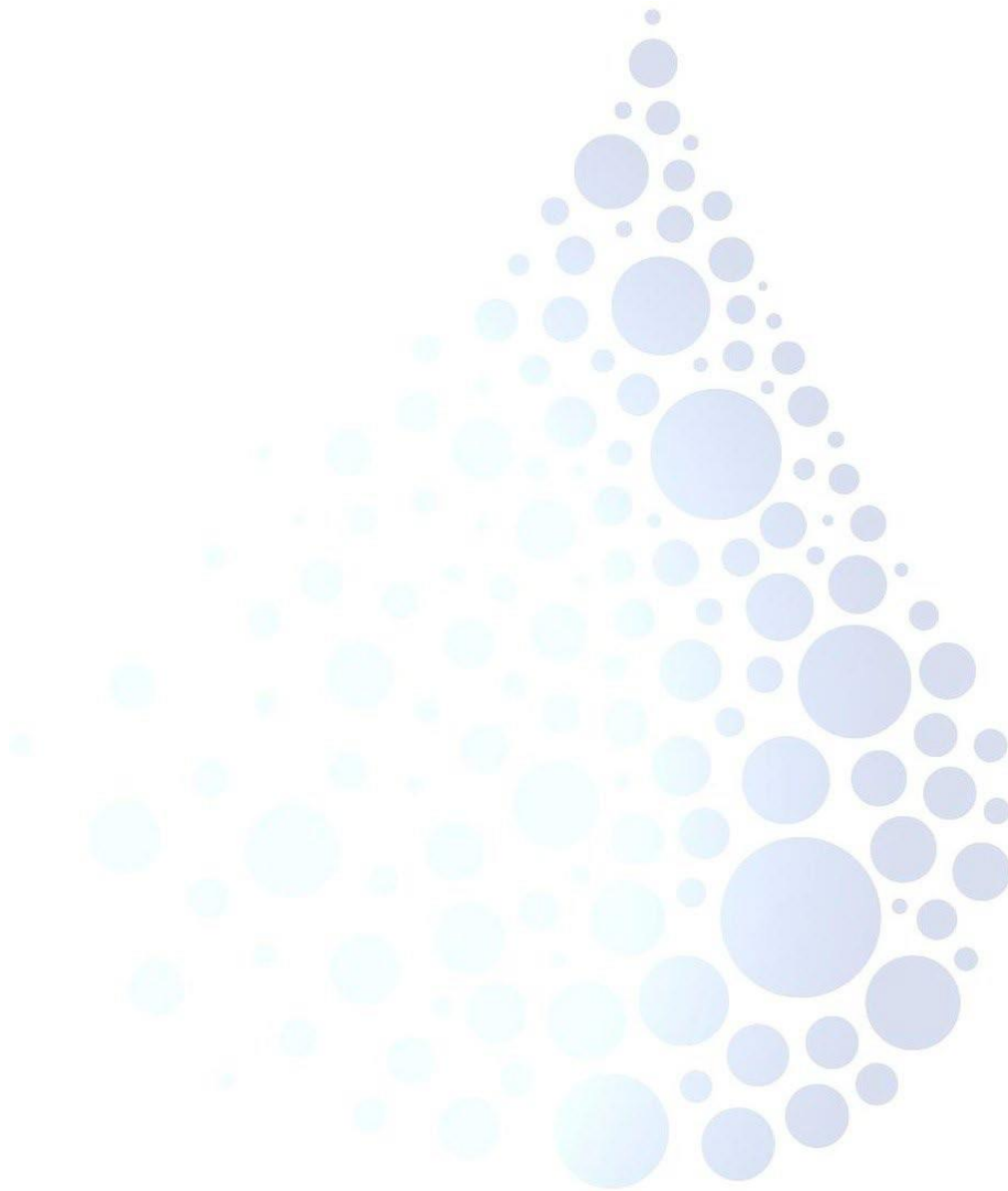
## References

- (1) Siemens 2021. Data sheet for SINAMICS Power module PM240-2, (2021).
- (2) WORLDWIDE ELECTRIC 2019. MOTOR DATA SHEET: PEWWE40-12-364T (2019).



**CCR**

Crystal Clearwater Resources



## Appendix 1- Water Analytical Results

Client Name: Client			Project Number: HP21-CCR-02-00			Date: January 21st, 2022			Rev #: A		
Analysis	Date (dd-mm-yy)	Time	Location	ID	Conductivity from inline Analyzer (µS/cm)	Density (g/cm3)	pH	Total Dissolved Solids (mg/L)	Total Dissolved Solids - Calculated Method (µEq for Focused Summation) (mg/L)	Conductivity (mS/cm)	Total Chloride (mg/L)
General Water Quality (by APEX Environmental)											
	8-Nov-21	12:56:00 PM	Distillate	1K08013-01	2738		8.22	92	166.5		
	10-Nov-21	12:05:00 PM	Distillate	1K10005-01	4304		8.19	720	678.2		
	16-Nov-21	11:25:00 AM	Distillate	1K16007-01	3355		8.12	52	163.4		
Complete Water Analysis (by Oilfield Labs of America)											
	30-Oct-21	N/A	Freshwater Feed	1211101001		1.0047	7.7	5383.5	5991.4	NR	6
	30-Oct-21	N/A	Freshwater Feed	1211101002		1.0036	8.08	3685.8	4727.5	NR	0
	30-Oct-21	N/A	Vacuum Vent	1211101003		1.0018	4.731	939.4	767.2	NR	240
	3-Nov-21	N/A	PWFEEED PRE	1211103096		1.0832	6.538	126800	126649.3	NR	169
	3-Nov-21	N/A	PWFEEED P	1211103097		1.0835	6.648	127224.1	127093.7	NR	161
	4-Nov-21	10:30:00 PM	Evap Loop	1211105112	2741	1.0837	6.821	127594.9	127396.8	NR	250
	4-Nov-21	10:30:00 PM	Feed	1211105118	170010	1.058	6.604	87850.1	87700.3	NR	144
	4-Nov-21	10:30:00 PM	Distillate	1211105115	555	1.0017	9.711	745.8	566.3	NR	0
	5-Nov-21	12:44:00 PM	Distillate	1211105116	472	1.0014	9.042	367.7	241.3	NR	0
	5-Nov-21	8:51:00 AM	Evap Loop	1211105113	154870	1.0969	6.372	148083.2	147794.8	NR	356
	5-Nov-21	1:42:00 PM	Evap Loop	1211105114	156726	1.0933	6.354	142444.0	142160.6	NR	375
	5-Nov-21	1:34:00 PM	Distillate	1211105117	457	1.0014	9.059	367	241.5	NR	0
	6-Nov-21	2:45:00 AM	Evap Loop	1211108045	3092	1.1126	6.1	172341.5	172037.5	201	359
	6-Nov-21	2:45:00 AM	Distillate	1211108001	382	1.0015	8.774	527.3	427.4	3.966	0
	6-Nov-21	6:00:00 AM	Evap Loop	1211108046	180139	1.1175	6	179932.7	179645.2	205.8	347
	6-Nov-21	6:00:00 AM	Distillate	1211108002	382	1.0027	8.771	2330.3	2234.5	3.924	0
	6-Nov-21	12:34:00 PM	Feed	1211108048	164309	1.0652	6.643	99023.7	98836.5	137.1	187
	6-Nov-21	12:32:00 PM	Evap Loop	1211108047	132574	1.1046	6.196	159955.8	159715.8	193.8	295
	6-Nov-21	12:24:00 PM	Distillate	1211108003	391	1.0017	8.771	788.9	694.3	3.918	0
	6-Nov-21	6:15:00 PM	Evap Loop	1211108151	126553	1.0988	6.094	151017.4	150785.3	186.4	286
	6-Nov-21	6:10:00 PM	Distillate	1211108143	382	1.0016	8.779	699	584.7	3.879	0
	6-Nov-21	2:09:00 PM	Evap Loop	1211108152	160161	1.1009	6.012	154246.6	154028.7	188.9	284
	6-Nov-21	2:11:00 PM	Distillate	1211108144	371	1.0014	8.846	394	248.4	3.704	0
	8-Nov-21	3:00:00 AM	Evap Loop	1211108153	767671	1.1193	5.722	182709.4	182426.1	207.3	355
	8-Nov-21	3:00:00 AM	Distillate	1211108145	349	1.0014	8.764	348.8	259.1	3.333	0
	8-Nov-21	11:52:00 AM	Feed	1211108155	121883	1.058	6.531	87909.6	87779.7	125.3	134
	8-Nov-21	11:45:00 AM	Evap Loop	1211108154	152837	1.0974	6.027	148859.7	148661.0	185.6	267
	8-Nov-21	11:45:00 AM	Distillate	1211108146	600	1.0015	8.726	415.9	314.5	6.344	0
	10-Nov-21	8:40:00 AM	Air Recooler Makeup	1211115180		1.0047	8.23	5489.7	5477.8	9.751	0
	10-Nov-21	12:44:00 PM	Feed	1211115182	98136	1.0502	6.72	75743.4	75665.6	107.3	55
	10-Nov-21	12:24:00 PM	Evap Loop	1211115181	152486	1.1038	6.11	158632.4	158554.2	184.9	144
	10-Nov-21	12:15:00 PM	Distillate	1211115172	1320	1.0019	8.4	1070	928.9	1.425	45
	10-Nov-21	5:09:00 PM	Distillate	1211115173	333730	1.0015	8.3	486	329.6	3.79	58
	11-Nov-21	11:00:00 PM	Feed	1211115184	128815	1.0744	6.53	113171.3	113043.8	150.1	114
	11-Nov-21	11:00:00 PM	Evap Loop	1211115183	601027	1.1037	6.01	158566.4	158382.7	200.5	207
	11-Nov-21	11:00:00 PM	Distillate	1211115174	418	1.0015	8.4	509	339.5	3.984	61
	12-Nov-21	8:30:00 PM	Distillate	1211115175	340	1.0016	8.3	648	425.2	3.364	101
	13-Nov-21	4:30:00 AM	Feed	1211115186	3750	1.0831	6.63	126738.2	126442.2	131.3	70
	13-Nov-21	4:30:00 AM	Evap Loop	1211115185	306149	1.0831	6.63	126636.2	126493.2	130.9	71
	13-Nov-21	4:30:00 AM	Distillate	1211115176	349	1.0016	8.4	581	381.0	3.579	74
	13-Nov-21	2:00:00 PM	Distillate	1211116267	350	1.0014	8.5	349	250.7	3.207	0
	13-Nov-21	4:35:00 PM	Distillate	1211116268	340	1.0013	8.4	242	167.9	3.116	0
	13-Nov-21	8:30:00 PM	Evap Loop	1211116265	263151	1.1047	5.903	160012.8	159720.9	200.3	341
	13-Nov-21	8:30:00 PM	Distillate	1211116269	317	1.0013	8.4	233	152.2	3.188	0
	16-Nov-21	4:00:00 AM	Feed Tank	1211116266		1.0051	6.833	5985.2	5982.2	10.6	17
	16-Nov-21	3:24:00 PM	Feed	121118098	132771	1.0692	6.505	105120.5	104976.4	140.1	138
	16-Nov-21	3:24:00 PM	Evap Loop	121118097	143410	1.1241	5.93	190088.6	189854.6	204.7	293
	16-Nov-21	3:24:00 PM	Distillate	121118093	22523	1.0017	7.901	836.2	654.9	0.2467	76
	17-Nov-21	12:50:00 PM	Distillate	121118094	2227	1.0015	7.79	439.3	311.6	0.2543	63
	18-Nov-21	7:50:00 PM	Distillate	121118095	427	1.0015	8.22	552.5	402.0	0.3641	43
	17-Nov-21	6:30:00 PM	Freshwater	121118099		1.0037	7.869	3832.1	3773.6	6.807	20
	4-Dec-21	9:45:00 AM	Feed	1211207220	409	1.0871	6.472	132859.5	132253.3	NR	619
	4-Dec-21	9:45:00 AM	Evap Loop	1211207218	409	1.1845	5.404	283425.4	282319.8	NR	1220
	4-Dec-21	9:45:00 AM	Distillate	1211207219	409	1.0015	8.506	545.9	372.3	NR	96
	4-Dec-21	3:30:00 PM	Distillate	1211207221	300	1.0014	8.925	384.4	222.6	NR	93
	5-Dec-21	3:30:00 PM	Distillate	1211207222	315	1.0015	8.045	521.8	319.8	NR	115
	6-Dec-21	8:00:00 AM	Distillate	1211207217	306	1.0015	8.311	497	297.5	NR	121
TOG											
	3-Nov-21	N/A	Feed	1211103099							231
TPH											
	3-Nov-21	N/A	Feed	1211103098							210
	4-Nov-21	10:30:00 PM	Evap Loop	1211105119							7
	4-Nov-21	10:30:00 PM	Distillate	1211105120							14
	6-Nov-21	2:45:00 AM	Distillate	1211108004							12
	6-Nov-21	6:00:00 AM	Distillate	1211108005	382						13
	6-Nov-21	12:34:00 PM	Feed	1211108050							29
	6-Nov-21	12:32:00 PM	Evap Loop	1211108049							15
	6-Nov-21	12:24:00 PM	Distillate	1211108006							13
	6-Nov-21	2:09:00 PM	Evap Loop	1211108156							5
	6-Nov-21	2:11:00 PM	Distillate	1211108147							5
	8-Nov-21	3:00:00 AM	Evap Loop	1211108157							14
	8-Nov-21	3:00:00 AM	Distillate	1211108148							6
	8-Nov-21	11:52:00 AM	Feed	1211108159							21
	8-Nov-21	11:45:00 AM	Evap Loop	1211108158							10
	8-Nov-21	11:45:00 AM	Distillate	1211108149							13
	10-Nov-21	12:44:00 PM	Feed	1211115188							34
	10-Nov-21	12:24:00 PM	Evap Loop	1211115187							22
	10-Nov-21	12:15:00 PM	Distillate	1211115177							11



# Water Analytical Results: Cations & Anions

Client Name: Client

Rev #:

A

Project Number: HP21-CBR-02-00

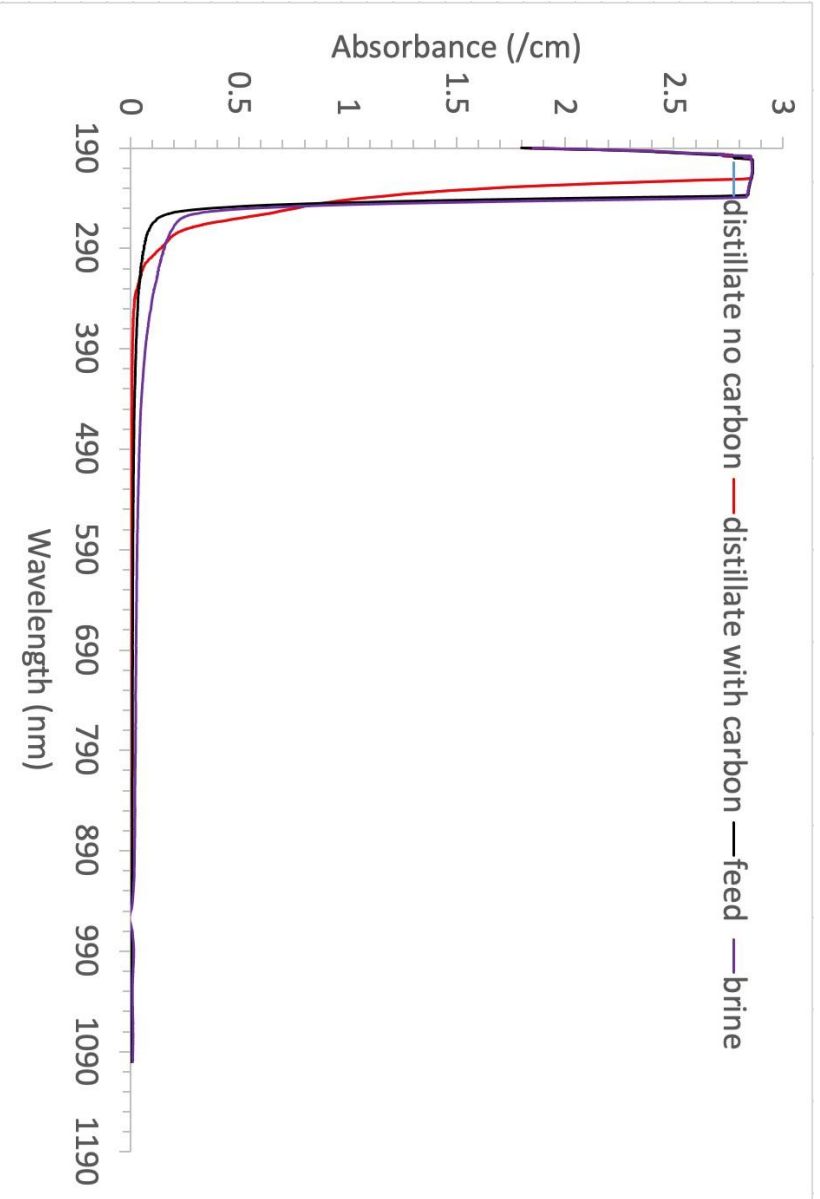
Date: January 21st, 2022

Analysis (General Water Quality by AP/EK Environmental)	Date (Reference)	Time	Location	Lab Name	ID	[Fe] (mg/L)	Manganese [Mn] (mg/L)	Sodium [Na] (mg/L)	Calcium [Ca] (mg/L)	Magnesium [Mg] (mg/L)	Potassium [K] (mg/L)	Barium [Ba] (mg/L)	Sulfate [S] (mg/L)	Aluminum [Al] (mg/L)	Lithium [Li] (mg/L)	Zinc [Zn] (mg/L)	Lead [Pb] (mg/L)	Carbonate [HCO <sub>3</sub> -] (mg/L)	Sulfate [SO <sub>4</sub> -2] (mg/L)	Chloride [Cl-] (mg/L)	Cyanide (mg/L)	Corrected Chloride [Cl-] (mg/L)	Borohydroxide [B(OH) <sub>3</sub> ] (mg/L)	Fluoride [F-] (mg/L)	Phosphate [PO <sub>4</sub> -3] (mg/L)	[Si] (mg/L)
8-Nov-21	12:36:00 PM		Ditch/Line	HK0813-501		24.8	3.66	190	24.6	1.35	202.461	15.12	0.020711	15.37	0.020379	0.073548	0.032255	0.089905	0.02026	160	42.6		14.4068			
10-Nov-21	12:05:00 PM		Ditch/Line	HK-0005-01		19.0	24.6	190	24.6	1.35	15.12	0.020711	15.37	0.020379	0.073548	0.032255	0.089905	0.02026	160	42.6		14.4068				
16-Nov-21	11:28:00 AM		Ditch/Line	HK-6007-201		18.8	7.23	15.35											137	19	34.8					
30-Oct-21	N/A		Freshwater Feed	1211101001		2.064	0.039576	974.541	77.368	150.086	202.461	0.142192	22.289	0.073548	0.032255	0.089905	0.02026	160	42.6		14.4068					
30-Oct-21	N/A		Freshwater Feed	1211101002		0.17546	0.025744	662.277	75.173	133.869	15.12	0.020711	15.37	0.020379	0.073548	0.032255	0.089905	0.02026	160	42.6		14.4068				
30-Oct-21	N/A		Vacuum Vent	1211101003		118271	16.223	80.14	67.604	67.604	4.164	0.029974	0.294666	2.731	0.026411	2.269	0.02026	160	42.6		14.4068					
3-Nov-21	N/A		PVT/FEED P/E	1211102986		12.626	0.02023	4069.8	5554.64	622.015	712.54	8.12	1467.24	0	33.37	0.132646	0.027242	113.46	130	77.900		332.6898	2.925868	0.074748	1.934	13.585
4-Nov-21	N/A		PVT/FEED	1211102987		3.699	0.04001	4052.4	6062.92	668.232	776.526	8.28	1462.38	0	35.653	0.133787	0.026358	79.5	110	77.900		334.8182	2.919683	1.138	16.301	12.353
4-Nov-21	10:30:00 PM		Feed	1211103112		1.059	0.04001	4170.1	5101.13	664.919	664.919	1.179	990.144	0	35.246	0.132646	0.027242	113.46	130	77.900		334.8182	2.919683	1.138	16.301	12.353
4-Nov-21	10:30:00 PM		Feed	1211103112		0.865258	0.31419	2729.9	3614.97	596.635	474.686	0.44724	990.144	0	35.246	0.132646	0.027242	113.46	130	77.900		334.8182	2.919683	1.138	16.301	12.353
4-Nov-21	10:30:00 PM		Ditch/Line	1211103115		0	0.039712	0	2.744	1.152	0.936	0.026364	0.1317	0	0.34372	0.026364	0.044433	4417.4	210	0	20					
5-Nov-21	13:40:00 PM		Feed	1211103115		0	0.039712	0	2.744	1.152	0.936	0.026364	0.1317	0	0.34372	0.026364	0.044433	4417.4	210	0	20					
5-Nov-21	13:40:00 PM		Feed	1211103115		1.88	0.034526	4828.5	6222.78	696.803	774.116	1.102	1725.6	2.74	34.333	0.131452	0.044433	207.78	419	68.000		384.9343	3.918473	1.241	18.101	13.209
5-Nov-21	13:40:00 PM		Feed	1211103115		1.469	0.034526	4870.58	6222.78	696.803	774.116	1.102	1725.6	2.74	34.333	0.131452	0.044433	207.78	419	68.000		384.9343	3.918473	1.241	18.101	13.209
5-Nov-21	13:40:00 PM		Feed	1211103115		0	0.034526	4870.58	6222.78	696.803	774.116	1.102	1725.6	2.74	34.333	0.131452	0.044433	207.78	419	68.000		384.9343	3.918473	1.241	18.101	13.209
5-Nov-21	2:45:00 AM		Feed	1211103095		1	1.101	15.098	1718.41	1186.43	832.947	2.88	1903.31	0	51.48	0.14315	0.179648	82.96	310	107.000		472.1759	38.89122	0.87345	18.482	0.072895
5-Nov-21	6:00:00 AM		Feed	1211103095		1.117	0.003689	185.831	3.273	1.919	0	1.302	0	0.03097	0.020362	0.020361	240.34	30	90	0	0					
5-Nov-21	6:00:00 AM		Feed	1211103095		0.000301	0.004165	798.148	8.89	1.464	1.442	0.03363	0.733128	0	28.614	0.119102	0.143579	29.68	100	110.000		461.5835	37.20287	0.819223	17.65	0.189888
5-Nov-21	12:30:00 PM		Feed	1211103096		1.034	1.008	42870.6	6846.13	1100.53	774.94	6.371	1174.12	0	46.872	0.142672	0.022629	75.84	330	380		265.6368	22.048654	1.098	10.303	18.275
5-Nov-21	12:30:00 PM		Feed	1211103096		0.000301	0.004165	798.148	8.89	1.464	1.442	0.03363	0.733128	0	28.614	0.119102	0.143579	29.68	100	110.000		461.5835	37.20287	0.819223	17.65	0.189888
5-Nov-21	12:30:00 PM		Feed	1211103096		1.034	1.008	42870.6	6846.13	1100.53	774.94	6.371	1174.12	0	46.872	0.142672	0.022629	75.84	330	380		265.6368	22.048654	1.098	10.303	18.275
5-Nov-21	12:30:00 PM		Feed	1211103096		0.000301	0.004165	798.148	8.89	1.464	1.442	0.03363	0.733128	0	28.614	0.119102	0.143579	29.68	100	110.000		461.5835	37.20287	0.819223	17.65	0.189888
6-Nov-21	6:15:00 PM		Ditch/Line	1211101003		0.040012	0.000102	4170.1	4127.33	307.913	4127.33	0.16327	1.171	0	44.49	0.143785	0	68.32	100	98.200		434.2248	38.212154	1.036	18.079	0.044718
6-Nov-21	6:15:00 PM		Ditch/Line	1211101003		0.040012	0.000102	4170.1	4127.33	307.913	4127.33	0.16327	1.171	0	44.49	0.143785	0	68.32	100	98.200		434.2248	38.212154	1.036	18.079	0.044718
6-Nov-21	2:09:00 PM		Ditch/Line	1211101015		10.3	0.390106	44081.6	2.368	1.206	10.729	0.021987	0.822791	0	46.379	0.145042	0.081154	47.28	150	100		0	0	0	0	0
6-Nov-21	2:10:00 PM		Ditch/Line	1211101015		0	0.007188	0	11.802	1.206	10.729	0.021987	0.822791	0	46.379	0.145042	0.081154	47.28	150	100		0	0	0	0	0
8-Nov-21	3:00:00 AM		Ditch/Line	1211103855		8.9	1.153	51860	7679.74	1200.41	818.668	4.303	1904.4	0	54.282	0.141582	0	51.24	320	118.000		501.00037	40.291383	1.938	18.812	0
8-Nov-21	3:00:00 AM		Ditch/Line	1211103855		0	0.003179	0	1.91	1.177	1.177	0	1.91	0	0.018984	0	2.1838	10	100	100		1.5322	0	0.202593	0	0
8-Nov-21	11:45:00 AM		Feed	1211103855		10	0.59076	24596.1	3772.63	621.251	384.425	6.983	1044.47	0	24.607	0.100689	0.0132	108.54	660	95.000		252.0719	25.052474	3.299	11.697	11.697
8-Nov-21	11:45:00 AM		Feed	1211103855		9.2	0.02916	41156	6442.07	1007.23	670.282	1.974	1306.68	0	43.941	0.143366	0.086734	38.6	570	96.000		451.5275	40.940244	2.463	19.115	19.115
8-Nov-21	11:45:00 AM		Feed	1211103855		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
8-Nov-21	11:45:00 AM		Feed	1211103855		0.2	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932	0	0.051579	247.66	10	130		1.3213	0	0.079785	0	0
10-Nov-21	12:40:00 PM		Feed	1211115162		0.5	0.020448	0	7.284	2.279	12.4	0.016019	12.784	0	18.932											



50e-21 3:30:00 PM Deline 1217207222 0.188419 0.00454 76.335 4.79 1.001 0.76393 0.027612 0.426866 0.005053 0.001625 0.006926 0 204.96 30 80 1.0293 0.029524 2.236 0.013798  
60e-21 8:00:00 AM Deline 1217207217 0.136311 0.004724 101.133 7.376 1.627 1.08 0.028127 0.346812 0.002082 0.006589 0.010995 0 189.1 0 70 1.4482 0.1566288 0.781568 0.07313

## Appendix 2- Water Analytical Results provided by NMPWRC



	feed	brine	distillate (no GAC)	distillate (post GAC)
Cond. ( $\mu\text{S}/\text{cm}$ )			347.6-348.1	1025
TDS (mg/L)	130,020 $\pm$ 3,048	212,050 $\pm$ 4,766		
TOC (mg/L)	72.33	103.3	43.29	77.26
pH	6.3-6.33	5.25	8.69-8.73	9.51-9.53
NH <sub>3</sub> -N (mg/L)	610	968	46.4	40.6

Note: TDS was measured by EPA standard evaporation methods

The precise cause of the distillate conductivity increasing during carbon polishing is not known, although one possible reason is that residue present on the inside of the drums, or on the activated carbon pellets themselves was mobilized during flow-through. The polishing step was a primitive process using a small centrifugal pump and two 50-gallon drums. Polishing operations for a full-scale deployment plant would be properly designed for throughput and residence time.

	m0, g	V, ml	m	TDS, mg/L		
feed PW	45.0478	5		45.6892	128280	feed, mg/L
	51.113	5		51.7542	128240	ave 130020
	31.7568	5		32.4245	133540	std 3048.475
Brine	28.9481	5		29.9915	208680	brine, mg/L
	31.0367	5		31.0894	10540	ave 212050
	30.5226	5		31.5997	215420	std 4765.8997