PILOT TESTING OF A MEMBRANE SYSTEM FOR POST-COMBUSTION CO_2 CAPTURE

Final Report
Report Period: October 1, 2010 – September 30, 2015

by

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ABSTRACT

This final report summarizes work conducted for the U.S. Department of Energy, National Energy Technology Laboratory (DOE) to scale up an efficient post-combustion CO_2 capture membrane process to the small pilot test stage (award number DE-FE0005795). The primary goal of this research program was to design, fabricate, and operate a membrane CO_2 capture system to treat coal-derived flue gas containing 20 tonnes CO_2/day (20 TPD). Membrane Technology and Research (MTR) conducted this project in collaboration with Babcock and Wilcox (B&W), the Electric Power Research Institute (EPRI), WorleyParsons (WP), the Illinois Sustainable Technology Center (ISTC), Enerkem (EK), and the National Carbon Capture Center (NCCC). In addition to the small pilot design, build and slipstream testing at NCCC, other project efforts included laboratory membrane and module development at MTR, validation field testing on a 1 TPD membrane system at NCCC, boiler modeling and testing at B&W, a techno-economic analysis (TEA) by EPRI/WP, a case study of the membrane technology applied to a ~20 MW_e power plant by ISTC, and an industrial CO_2 capture test at an Enerkem waste-to-biofuel facility.

The 20 TPD small pilot membrane system built in this project successfully completed over 1,000 hours of operation treating flue gas at NCCC. The Polaris™ membranes used on this system demonstrated stable performance, and when combined with over 10,000 hours of operation at NCCC on a 1 TPD system, the risk associated with uncertainty in the durability of post-combustion capture membranes has been greatly reduced. Moreover, next-generation Polaris membranes with higher performance and lower cost were validation tested on the 1 TPD system. The 20 TPD system also demonstrated successful operation of a new low-pressure-drop sweep module that will reduce parasitic energy losses at full scale by as much as 10 MW_e.

In modeling and pilot boiler testing, B&W confirmed the viability of CO_2 recycle to the boiler as envisioned in the MTR process design. The impact of this CO_2 recycle on boiler efficiency was quantified and incorporated into a TEA of the membrane capture process applied to a full-scale power plant. As with previous studies, the TEA showed the membrane process to be lower cost than the conventional solvent capture process even at 90% CO_2 capture. A sensitivity study indicates that the membrane capture cost decreases significantly if the 90% capture requirement is relaxed. Depending on the process design, a minimum capture cost is achieved at 30-60% capture, values that would meet proposed CO_2 emission regulations for coal-fired power plants.

In summary, this project has successfully advanced the MTR membrane capture process through small pilot testing (technology readiness level 6). The technology is ready for future scale-up to the 10 MW_e size.

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ABSA	Alberta Boiler Safety Association
ACFM	Actual cubic feet per minute
AERF	Advanced Energy Research Facility in Edmonton, AB
AI-EES	Alberta Innovates-Energy and Environmental Solutions
APS	Arizona Public Services, Cholla
ATR	autothermal reformer
BACT	Best Available Control Technology
B&W	Babcock and Wilcox
BBS	Bituminous Baseline study
BOP	Balance of plant
BSR	boiler stoichiometric ratio
CFD	Computational fluid dynamics
CFR	Code of Federal Regulations
CO_2	Carbon dioxide
COE	Cost of electricity
CoE	City of Edmonton
COMO^{SM}	Combustion Model (B&W software)
CPU	compression and purification unit
CRN	Canadian registration number
CSA	Compliance Safety Accountability
DCC	Direct contact cooler
DOE	U.S. Department of Energy, National Energy Technology Laboratory
EAB	Enerkem Alberta Biofuels
EK	Enerkem
EOR	Enhanced oil recovery
EPA	Environmental Protection Agency
EPRI	Electric Power Research Institute
FAT	Factory Acceptance Test
FD	forced draft
FEGT	furnace exit gas temperature

FGD	flue gas desulfurization
Gen-1	first-generation
Gen-2	second-generation
Gen-3	third-generation
gpm	Gallons per minute
gpu	gas permeation unit
h	hours
HAZOP	Hazard and operability study
IGCC	Integrated Gasification Combined Cycle
ISTC	Illinois Sustainable Technology Center
kscfd	thousand standard cubic feet per day
kW$_e$	kilowatts electrical
kWh/tonne	kilowatt hours per tonne
lb	pound
lb/h	pound per hour
LOI	Combustible loss
m	meter
m^2	meters squared
m^3/h	cubic meters per hour
m^3/min	cubic meters per minute
MBtu	thousand British thermal units
MCC	Motor Control Center
MEA	monoethanolamine
MTR	Membrane Technology and Research, Inc.
MW	megawatt
MW$_e$	megawatt electrical
MWh	megawatt hour
MWh$_{net}$	megawatt hours net
η	boiler efficiency
NCCC	National Carbon Capture Center
NDIR	nondispersive infrared
NETL	National Energy Technology Laboratory

NGCC	Natural Gas Combined Cycle
NOAK	Nth of a kind
O&M	operating and maintenance
P	Permeability
PC	pulverized coal
PC4	Post-Combustion Carbon Capture Center
PCFG	pulverized coal flue gas
PLC	programmable logic controller
POTW	Publicly Owned Treatment Works
ppm	parts per million
PRB	Powder River Basin
psi	pounds per square inch
psia	pounds per square inch absolute
psig	pounds per square inch gauge
PSTU	Pilot Solvent Test Unit
R&D	research and development
RB	radiant boiler
RCRA	Resource Conservation and Recovery Act
RO	reverse osmosis
S	Solubility
SA	secondary air
SBS-II	Small Boiler Simulator II
SCFM	standard cubic feet per minute
SEGR	selective exhaust gas recycle
SS	stainless steel
SWUP	spiral-wound ultra-supercritical universal pressure
TEA	techno-economic analysis
TEG	triethylene glycol
tonne	1,000 kg
TOx	thermal oxidizer
TPD	tonnes of CO_2 per day
TRL	Technology Readiness Level

UBC	unburned carbon
WFGD/DFGD	Wet FGD/Dry FGD
WP	WorleyParsons
wt%	weight percent
$\alpha_{1/2}$	selectivity
°C	degrees Celsius
°F	degrees Fahrenheit
°K	degrees Kelvin
µg/mL	micro grams per milliliter
µm	microns

EXECUTIVE SUMMARY

Introduction

This final report describes work conducted for the U.S. Department of Energy, National Energy Technology Laboratory (DOE) to scale-up an efficient post-combustion CO_2 capture membrane process to the small pilot test stage (award number DE-FE0005795). The primary goal of this research program was to design, fabricate, and operate a membrane CO_2 capture system to treat coal-derived flue gas containing 20 tonnes CO_2/day [20 TPD] (corresponding to the amount of CO_2 generated by 1 MW_e of coal-fired power production). Membrane Technology and Research, Inc. (MTR) conducted this project in collaboration with Babcock and Wilcox (B&W), the Electric Power Research Institute (EPRI), WorleyParsons (WP), the Illinois Sustainable Technology Center (ISTC), Enerkem (EK), and the National Carbon Capture Center (NCCC). In addition to the small pilot design, build and slipstream testing at NCCC, other project efforts included laboratory membrane and module development at MTR, validation field testing on a 1 TPD membrane system at NCCC, boiler modeling and testing at B&W, a techno-economic analysis by EPRI, WP and MTR, a case study of the membrane technology applied to a ~20 MW_e power plant by ISTC, and an industrial CO_2 capture test at an Enerkem waste-to-biofuel facility. The work was conducted by the project partners from October 1, 2010 through September 30, 2015.

Carbon capture from power plant flue gas is difficult for all separation technologies, including membranes, because of the low partial pressure of CO_2 in flue gas. In previous DOE-funded work (DE-NT43085 and DE-FE0005312), MTR made two innovations to address the challenges of CO_2 capture from power plant flue gas with membranes:

- New membranes with CO_2 permeances approximately tenfold higher than commercial CO_2-selective membranes were developed. The high permeance of these new membranes – designated Polaris™ – greatly reduces the required membrane area, footprint, and capital cost of a membrane CO_2 capture system.

- A membrane selective recycle process was developed. This patented process uses combustion air as a sweep stream to generate driving force for transmembrane CO_2 transport. The separated CO_2 is recycled to the boiler with air. This design increases the concentration of CO_2 in flue gas, which reduces the energy and capital required for subsequent capture.

These innovations led to a focused effort by MTR, with DOE support, to move the Polaris membrane technology toward commercial readiness. Figure ES 1 shows the development timeline for the MTR CO_2 capture process. An initial feasibility project starting in 2007 (DE-NT43085) first proposed and modeled a selective recycle membrane process. During this project, the Polaris membrane configuration was also first conceived and tested in the lab. This work led to a follow-on project (DE-FE0005312) that included the first test of membrane modules with coal-fired flue gas at the Arizona Public Services (APS) Cholla plant in 2010. The APS Cholla test utilized a 50 kW_e or 1 TPD membrane unit that housed 8-inch diameter spiral-wound Polaris modules. This test demonstrated that Polaris membranes were robust enough to survive months of coal flue gas treatment without showing degradation, and it was the first field validation of air sweep module performance. After completion of the Cholla project, the current program (DE-

FE0005795) was initiated with a primary goal of scaling up the MTR capture approach to a 20 TPD small pilot test system. This report describes the key findings from the project, and includes results from testing of a new low-pressure-drop, plate-and-frame sweep module developed concurrently under a separate DOE program (DE-FE0007553). In summary, these activities over the past 8 years have brought the MTR post-combustion capture membrane technology from concept (TRL 2/3) through small pilot validation (TRL 6).

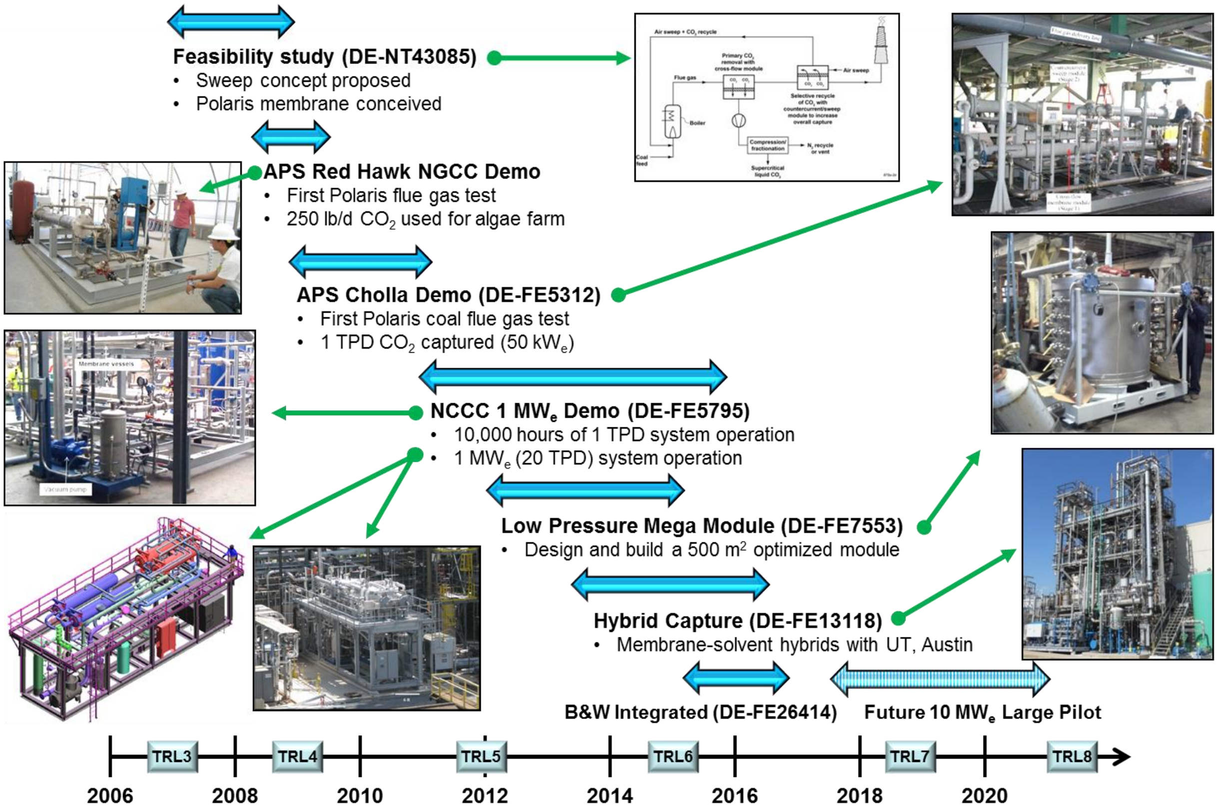


Figure ES 1. CO_2 capture development timeline for the MTR Polaris membrane process. The approximate Technology Readiness Level (TRL) is shown along the x-axis.

Membrane Performance Improvements and Cost Reduction

Polaris^{TM} membranes developed by MTR in previous DOE-funded work set the performance standard for post-combustion CO_2 capture membranes. With a CO_2 permeance of 1,000 gpu and CO_2/N_2 selectivity of 50, this designer membrane was an order of magnitude more permeable than conventional CO_2 separation membranes used for natural gas treatment. However, there are compelling reasons to strive for higher membrane CO_2 permeance: doubling the CO_2 permeance will roughly halve the required membrane area, and thus reduce the capital cost and footprint of a membrane CO_2 capture system. In fact, prior techno-economic sensitivity studies determined that increasing the CO_2 permeance and reducing the production cost of Polaris membranes was important to reach the DOE's capture cost target of <\$40/tonne CO_2. With this in mind, a goal was set at the start of this project to achieve at least a 50% improvement in CO_2 permeance without a loss of selectivity compared to the base line or Generation 1 (Gen-1) Polaris membranes.

Figure ES 2 shows the improvements made in Polaris membrane performance during this project in the form of a tradeoff plot, where CO_2/N_2 selectivity is plotted against CO_2 permeance. Over time, the Polaris membrane performance has steadily improved, particularly by increasing CO_2 permeance (data points move to the right on the figure). A Gen-2 version of the Polaris membrane was scaled up to pilot production. This membrane offers a CO_2 permeance of about 1,700 gpu, which exceeds the original project target. These Gen-2 membranes have been made on commercial roll-to-roll casting and coating equipment, fabricated into full-sized modules, and validated in field testing at NCCC. By the end of the project, advanced membranes with 3 times the CO_2 permeance of the Gen-1 Polaris membrane were being produced in the laboratory (3,000 gpu versus 1,000 gpu). Such high performance membranes make the case for membrane-based CO_2 capture compelling. Based on technical limitations and diminishing economic returns realized at even higher performance values (described later in this report), we believe these advanced 3,000 gpu membranes are a practical target for use when the technology is ready for full-scale commercial deployment.

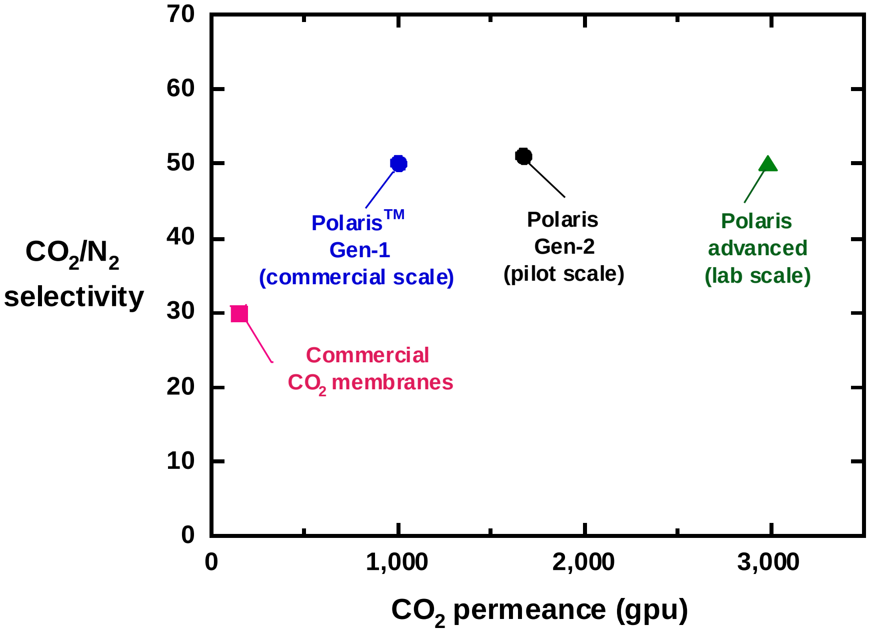


Figure ES 2. A CO_2/N_2 trade-off plot showing data for several generations of MTR Polaris membranes, compared with the properties of a good commercial natural gas membrane.

In addition to performance improvements, another key issue for the competitiveness of a membrane-based CO_2 capture system is the cost of the membranes and modules, which comprise a significant fraction of the capital cost of a large membrane system. At the start of this project, Gen-1 Polaris membrane was produced on a lab scale, and packaged into high-pressure gas separation modules where the total module cost was about \$500/m^2. Ultimately, economic analyses indicate a membrane module cost of \$50/m^2 is needed to meet DOE targets.

During this project, the Gen-1 Polaris membrane was scaled-up to commercial roll-to-roll production equipment. Twenty Polaris membrane rolls of 200 m length x 1 m width were used to produce the modules tested on the 1 and 20 TPD systems throughout the project. The Gen-1 membrane is now a fully commercial product and has been sold into several natural gas and refinery membrane applications. As part of the production scale-up process, a variety of membrane/module advanced manufacturing and design optimizations were used to reduce Polaris module costs from \$500/m^2 to \$200/m^2. This represents a 60% cost reduction, which meets the project target improvement. A detailed pathway to the ultimate \$50/m^2 cost target that is achievable by the post-2020 commercialization timeframe using advanced manufacturing concepts was also developed and is discussed in Section 2.3.

Membrane Lifetime and Validation Testing at NCCC

To validate the membrane performance improvements described above, a 1 TPD bench-scale membrane test system previously used at the APS Cholla plant was refurbished and installed at the NCCC. This system, shown in Figure ES 3(a), was installed in early 2012 and operated until July 2015, allowing testing of full-scale (8-inch diameter) spiral-wound Polaris modules with real coal-derived flue gas.

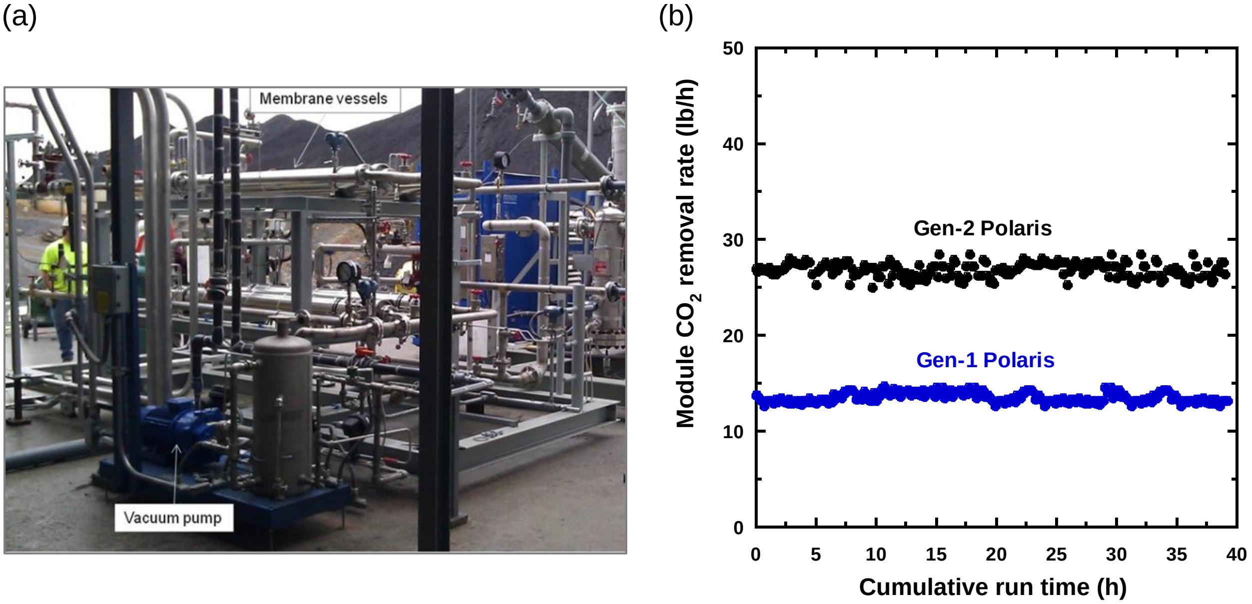


Figure ES 3. (a) Picture of the 1 TPD bench-scale test system at NCCC, and (b) module CO_2 capture rate as a function of operating time for two versions of the Polaris membrane.

Two types of tests were conducted on the 1 TPD system: membrane lifetime evaluation and validation testing of new membrane formulations. Lifetime testing is important because economic assessments must estimate a module replacement time, and prior to this program there was almost no data to base this estimate on. Commercial modules used in natural gas processing have average lifetimes of 3 to 5 years, so to be conservative, 3 years was taken as a first estimate for coal flue gas service. Over the course of this project, over 10,000 hours of operation was accumulated on the 1 TPD system showing stable module performance. Although not yet achieving 3 years of operation, this 1 TPD lifetime testing has greatly reduced the risk that Polaris modules will have degradation issues when treating coal flue gas.

The other important outcome from the 1 TPD system testing was validation of new generations of membrane treating real coal flue gas. Figure ES 3(b) shows an example of a short campaign where Gen-1 and Gen-2 Polaris modules were tested in parallel. The Gen-2 module removes about 70% more CO_2 than the Gen-1 module, confirming lab membrane tests.

20 TPD Small Pilot Design, Build and Operation

The primary objective of this project was design, construction and operation of a small pilot membrane system. Figure ES 4 shows a picture of the completed small pilot system installed at NCCC in July 2014. The two-story skid was pre-assembled at a fabricator and shipped to the site in two pieces corresponding to the two floors of the unit. In this way, the construction on-site was minimized. The top floor houses all of the membrane module vessels, which contained low-cost bundled spiral-wound Polaris modules. The bottom floor holds the rotating equipment (vacuum pump and feed compressor). The overall dimensions of the small pilot unit are 12 ft wide x 47 ft long x 23 ft tall. Design, construction and installation of the skid were completed as scheduled, on budget.

Figure ES 4. A picture of the MTR Polaris 20 TPD CO_2 capture system installed at NCCC.

Figure ES 5 shows a picture comparing the size and footprint of the MTR Polaris small pilot system with two different solvent capture systems of similar capacity installed at the NCCC's Post-Combustion Carbon Capture Center (PC4). PC4 is a slipstream test facility utilizing flue gas from Alabama Power Plant Gaston Unit 5, which is an 880 MW_e supercritical pulverized coal unit. The large 150 foot tall superstructure to the far right of the picture houses the NCCC Pilot Solvent Test Unit (PSTU) which tests various solvents for CO_2 capture on a scale of ~10 TPD. The tall structure in the center of the picture is an advanced solvent 20 TPD CO_2 capture system. In the far left of the picture is the MTR 20 TPD small pilot unit, shown previously in close up in Figure ES 4. The size and simplicity advantages of the membrane system are readily apparent.

Figure ES 5. A picture comparing the size and footprint of the MTR Polaris CO_2 capture system with two different solvent capture systems of similar capacity at NCCC.

The MTR 20 TPD system was operational during two NCCC test campaigns in the first half of 2015. The first campaign consisted of shakedown operations and parametric testing with bundled spiral-wound Polaris modules, while the second campaign focused on operation using new low-pressure-drop Polaris sweep modules. Figure ES 6 shows the CO_2 capture rate for the system as a function of run time during the second campaign in May and June 2015. The Polaris system achieved 90% CO_2 capture during parametric testing and consistently captured CO_2 at a rate of 85% or higher throughout the ~1,000 hour test campaign. Other than a few flue gas outages not related to the capture system, the membrane unit operated continuously and met design performance specifications.

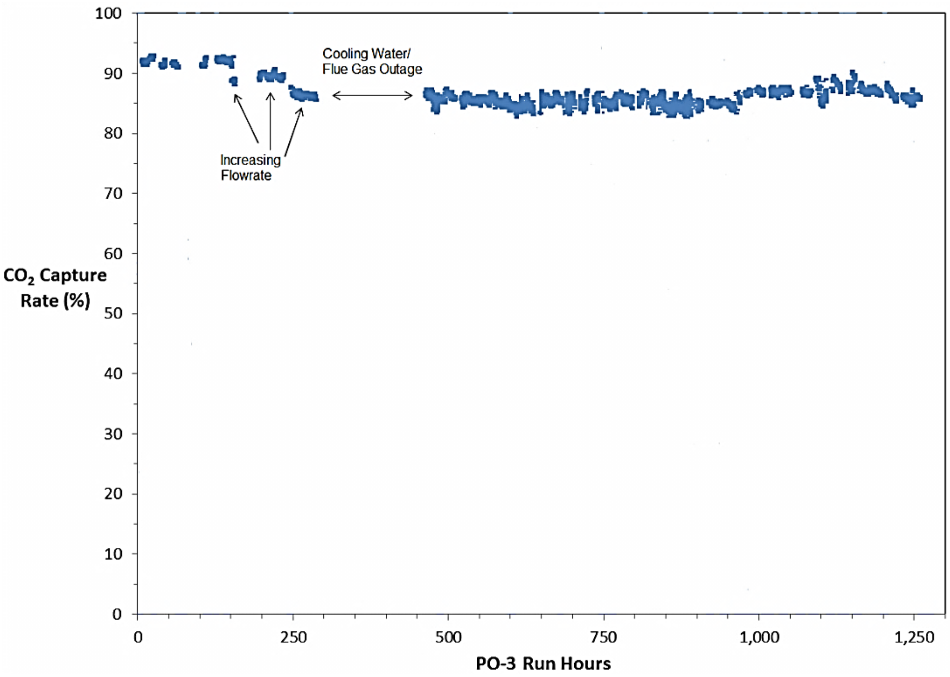


Figure ES 6. CO_2 capture rate as a function of operating time for the MTR 20 TPD system during NCCC campaign PO-3 (May – June 2015).

In addition to capture rate, another important performance metric for the 20 TPD small pilot system was the pressure drop through the membrane system. In particular, prior testing on the 1 TPD system at Cholla and NCCC showed that the pressure drop for the air sweep stream as it flows through spiral-wound sweep modules was unacceptably high. This high pressure drop would cause large parasitic energy losses on a full-scale system, reducing the efficiency of the selective recycle approach. To address this issue, a small prototype plate-and-frame module, designed to minimize sweep-side pressure drop, was built and lab tested in this project. Based on favorable results, a separate DOE program (DE-FE0007553) was initiated to optimize and scale-up these low pressure drop sweep modules. A prototype of these new sweep modules was incorporated into and tested on the 20 TPD unit at NCCC. Figure ES 7 compares the sweep side pressure drop measured on the 20 TPD system for the new plate-and-frame module with a conventional spiral-wound sweep module. Consistent with lab measurements, at the same flowrate, the pressure drop for the new module is more than four times lower than that of the spiral-wound sweep module. This difference would correspond to an energy savings of ~10 MW_e on a full scale system. Moreover, the pressure drop measured on the 20 TPD small pilot is even lower than the estimate used in the techno-economic analysis (~1.5 psi). This new plate-and-frame sweep module showed stable performance throughout the ~1,000 hour campaign, and will be the module configuration used for the air sweep step in future larger systems.

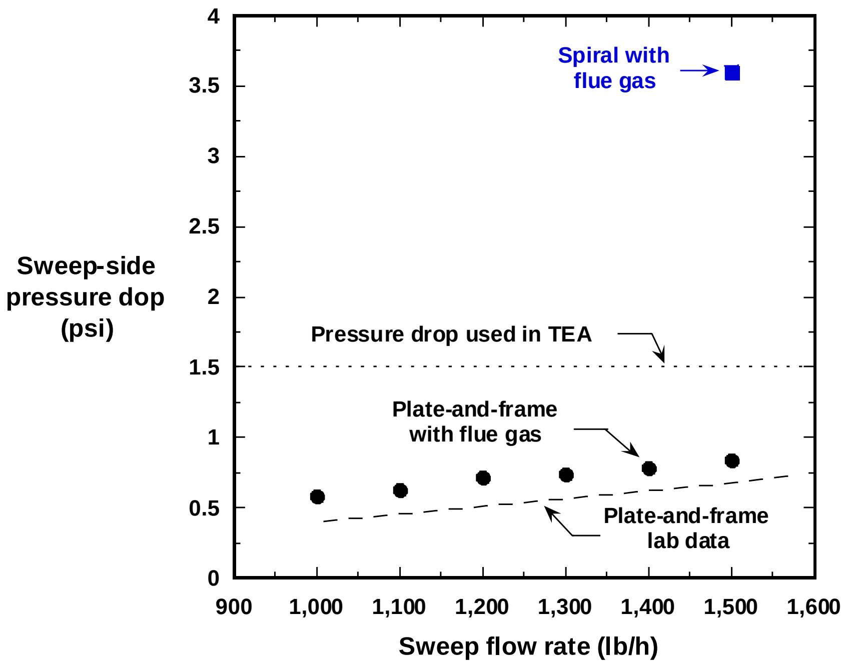


Figure ES 7. Comparison of sweep-side pressure drop of plate-and-frame and spiral-wound sweep modules during testing with flue gas on the 20 TPD system at NCCC. The dashed line represents the predicted plate-and-frame pressure drop based on lab data, while the dotted line represents the pressure drop used in the techno-economic analysis.

B&W Boiler Testing

One of the key differences between conventional post-combustion CO_2 capture approaches and the MTR membrane capture process is the selective recycle of CO_2 to the boiler used by the MTR process. Unlike other end-of-the-pipe capture technologies, the MTR process changes operating conditions in the boiler. By recycling CO_2, the boiler combustion air is diluted so that the concentration of oxygen is reduced compared to conventional operation. Quantifying the impact of recycled CO_2 on boiler performance, and its effect on the overall competitiveness of the MTR capture process was one of the objectives of this project. To this end, B&W conducted two studies during this project:

- a computational fluid dynamics analysis was performed using B&W's proprietary COMO^{SM} software to estimate flame stability, gas compositions, and heat distributions for various amounts of CO_2 recycle in two common pulverized coal (PC) boilers, and
- a pilot-scale test of boiler operation with CO_2-laden combustion air was conducted on B&W's Small Boiler Simulator (SBS-II) 0.6 MW_e research boiler.

The CFD modeling results suggested that CO_2 recycle in secondary air (SA) is feasible as a retrofit for either of the boiler configurations examined (a Carolina-type radiant boiler firing bituminous coal, and a spiral wound ultra-supercritical universal pressure boiler firing PRB coal) provided the boiler stoichiometry is maintained by increasing the secondary air flow. Under these conditions, the predicted combustion and heat transfer characteristics for CO_2-enriched air operation showed only a modest change from the baseline air-firing results. Based upon these favorable outcomes, a pilot-scale coal combustion and emissions performance evaluation with CO_2-enriched air was initiated.

B&W's SBS II was utilized for the pilot boiler testing. Two coals – a western sub-bituminous PRB and an eastern bituminous – were tested to study the effect of CO_2 addition to combustion air on boiler performance. To mimic CO_2 recycle by MTR's membrane process, CO_2 from a storage tank was added to the secondary air stream at SBS-II during experiments. The key outcomes from this boiler testing were:

- Stable and attached flames were observed for the combustion of both bituminous and sub-bituminous coals at windbox oxygen levels varying from 21% to 16% by volume (corresponding to different levels of CO_2 recycle in secondary air). This observation was confirmed by data generated using B&W's FlameDoctor statistical software, which showed better flame characteristics for CO_2-enriched combustion compared to conventional air firing.
- For the same burner configuration and vane settings, NO_x decreased as windbox oxygen was reduced from 21% to 16%. CO, on the other hand, showed a contrasting trend, increasing slightly from 16 ppm to 22 ppm. This general trend was observed for the combustion of the PRB and bituminous coals.
- Heat absorption in the boiler is shifted slightly from the furnace to the convective pass and the air heater for CO_2-enriched air combustion because of increased gas mass flow and lower flame temperature compared to normal air firing.
- Based on results from radiant furnace and convection pass heat absorption studies, pressure part modifications may not be required for CO_2-enriched air combustion at 18% windbox O_2. For this reason, in addition to good flame stability and mostly positive effects on emissions, burner windbox oxygen of 18% (corresponding to the base case MTR CO_2 recycle design) was judged to be optimum for retrofit cases.
- Boiler thermal efficiency decreased by approximately 2% at windbox O_2 of 18%. This corresponds to an overall plant efficiency loss of about 0.75% when selective CO_2 recycle is employed.

Figure ES 8 shows the estimated impact of CO_2 recycle on plant efficiency loss. As more CO_2 is recycled to the boiler, the windbox oxygen content decreases due to dilution. The plant efficiency loss increases in nearly a linear manner as O_2 concentration decreases. The 0.75% efficiency loss estimated from the B&W studies for the base case was incorporated into the TEA conducted in this project. This analysis indicates that at this relatively modest plant efficiency loss, there is a significant net benefit to the cost of capture for the membrane process by doing selective recycle of CO_2 to the boiler (particularly at high capture rates).

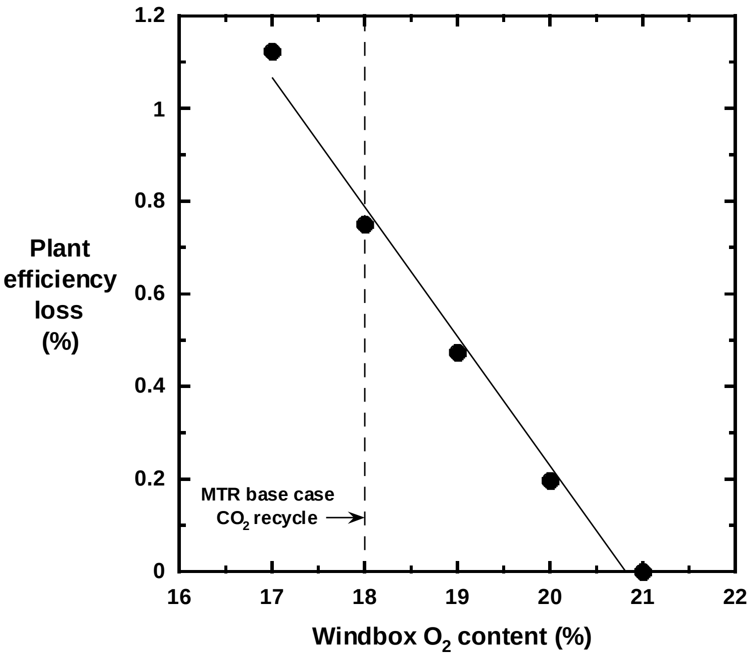


Figure ES 8. Estimated plant efficiency loss as a function of windbox oxygen content based on pilot boiler testing at B&W. The baseline amount of CO_2 recycle in MTR's process corresponds to a windbox oxygen content of 18%.

Techno-Economic Analysis

The MTR membrane capture process has been subjected to a number of prior techno-economic analyses (TEAs), including a DOE review of advanced capture processes (Current and Future Technologies for Power Generation with Post-Combustion Carbon Capture, DOE/NETL-2012/1557, also known as the "Pathways Study"). These previous studies indicate that the MTR membrane process offers a variety of advantages over conventional capture approaches and can reach DOE 90% CO_2 capture targets with future optimizations.

In this project, WP worked with MTR to estimate costs for a full-scale conceptual design of a supercritical coal-fired power plant with MTR's membrane process capturing 90% of the CO_2 emissions. EPRI used these costs to estimate the cost of electricity (COE) and CO_2 capture costs for this design, and to compare them with an absorption-based CO_2 capture process. In the following analysis, the base case for comparison is Case 11 from DOE's Bituminous Baseline Study (BBS) report, a supercritical pulverized-coal-fired plant without CO_2 capture. Two capture designs were compared to this case: a monoethanolamine (MEA)-based CO_2 capture process (BBS Case 12) and the MTR membrane process. The performance and costs for the Base and MEA cases are from the DOE's Updated Cost Report (August 2012) with all costs reported in June 2011 U.S. dollars (USD).

Table ES 1 provides a summary of the performance values for the three cases. All three plants are sized to produce approximately 550 MW_e net power output. Compared to the MEA case, the MTR membrane process produces significantly higher gross power output because the membrane system does not use steam, whereas MEA requires steam for solvent regeneration. On the other hand, the MTR process uses more auxiliary power than the MEA case because the membrane system requires electricity to drive blowers and vacuum pumps that provide driving force for CO_2 separation and to move gases through the system. Overall, the membrane process uses less energy for capture than the MEA case (358 kWh/tonne CO_2 vs. 382 kWh/tonne CO_2), and yields a slightly higher plant efficiency (28.6% vs 28.4%).

Table ES 1. Summary of EPRI/WP TEA Plant Performance Values.

Performance	BBS Case 11 (No CO_2 Capture)	BBS Case 12 (MEA)	MTR Membrane
Gross power output, MW_e	580.0	662.8	780.8
Auxiliary power requirements, MW_e (balance of plant)	30.4	65.5	186.5
Auxiliary power requirements, MW_e (capture system)	0	47.3	40.8
Auxiliary power requirements, MW_e (total)	30.4	112.8	227.3
Net power output, MW_e	559.0	550.0	553.5
Net plant HHV efficiency (%)	39.3	28.4	28.6
CO_2 generated (tonne/h)	440.3	608.1	608.4
Capture efficiency (%)	0	90.2	89.9
CO_2 emitted (tonne/h)	440.3	59.7	61.2
CO_2 captured (tonne/h)	0	548.4	547.2
CO_2 emissions (kg/MWh net)	801	109	111
Energy used per tonne CO_2 captured (kWh/tonne CO_2)	0	382	358

Table ES 2 compares the cost of electricity (COE) and CO_2 capture costs for the different cases. The cost of electricity is broken down into several categories, all of which experience a slightly smaller increase in cost relative to the no capture case for the membrane process compared to MEA. The largest cost difference between the MEA and membrane processes is the variable operating and maintenance (O&M) costs. The higher MEA variable O&M cost is largely attributable to the cost of replacing solvent. Overall, at 90% capture, the COE for the membrane case is $5/$MWh_{net}$ lower than the MEA process. Similarly, the cost of capture for the membrane process is about $4.5/tonne CO_2 (or 8%) lower than the MEA capture case.

Table ES 2. Cost Summary from the EPRI/WP TEA.

Cost	BBS Case 11 (No CO_2 Capture)	BBS Case 12 (MEA)	MTR Membrane
Cost of electricity components ($/$MWh_{net}$)	-	-	-
Fuel	25.5	35.3	35.1
Variable O&M	7.7	13.2	11.3
Fixed O&M	9.5	15.7	15.3
Capital	38.2	73.1	70.7
Total COE ($/$MWh_{net}$)	80.9	137.3	132.3
Increase in COE compared to Case 11 (%)	-	70	64
Cost of CO_2 captured ($/tonne)	-	56.5	52.0

In comparing these TEA results with prior studies, there is general agreement that the MTR membrane process offers advantages over conventional capture systems at 90% capture. MTR and WP used a conservative cost estimating philosophy in preparing this TEA. With the exception of the membrane modules, the balance of process equipment selected by WP (representing a significant fraction of the capital and operating costs), are commercial equipment that are available today in the required sizes. In that regard, the calculated COE and capture cost values presented here more closely relate to a first of a kind plant than an n^{th} of a kind plant. For comparison, the DOE Pathways Study examined MTR's capture system utilizing advanced compression/vacuum equipment and low risk technology financing and found capture costs of less than \$40/tonne CO_2.

Another variable that is expected to play a significant role in impacting the cost of capture is the capture rate. Membrane separation systems are well-known to be particularly cost-effective for bulk removal applications. While most of the CO_2 capture literature has focused on 90% capture, membranes are likely to be very attractive for lesser removal rates. Understanding the impact of capture rate on cost seems especially relevant given proposed CO_2 emissions regulations (such as the EPA Clean Power Plan) that amount to partial capture from coal-fired plants (for example, a proposed emissions limit of 1,400 lb CO_2/MWh amounts to <30% capture from an average coal-fired power plant).

With this fact in mind, MTR conducted a sensitivity study to examine the effect of CO_2 capture rate on the cost of capture for the membrane process. Figure ES9 shows these costs for two membrane design cases:

1. MTR's base case two-step selective recycle process (blue curve), and
2. A single step membrane process without recycle to the boiler (black curve).

Also shown in the figure is the data point at 90% capture for the base case design calculated in the WP/EPRI analysis (summarized in Table ES 2). The WP/EPRI capture cost is slightly higher than the sensitivity study value at 90% capture because the sensitivity study uses advanced compression and low risk financials.

The data in Figure ES 9 demonstrate two important points. First, the cost of capture for a membrane system is not constant with capture rate, and in fact, shows a minimum at <70% capture, the exact value of which depends on the process design. For the 2-step process with selective recycle, the minimum capture cost of ~\$35/tonne CO_2 occurs around 60% capture. At 90% capture, the cost for this process is on a steeply increasing trend. The second point is that for a single-step membrane process without selective recycle, the capture costs are much higher at 90% capture (>\$70/tonne). In fact, the difference between the cost curves for the two processes shows the benefit of selective recycle, which averages about \$20/tonne at capture rates >70%. However, at lower capture rates (for example, between 20 – 40%), the difference between the capture costs for the two processes is very small. Considering the simplicity of the one step process without recycle, this may be the preferred approach for a membrane capture system that meets proposed EPA CO_2 emissions limits.

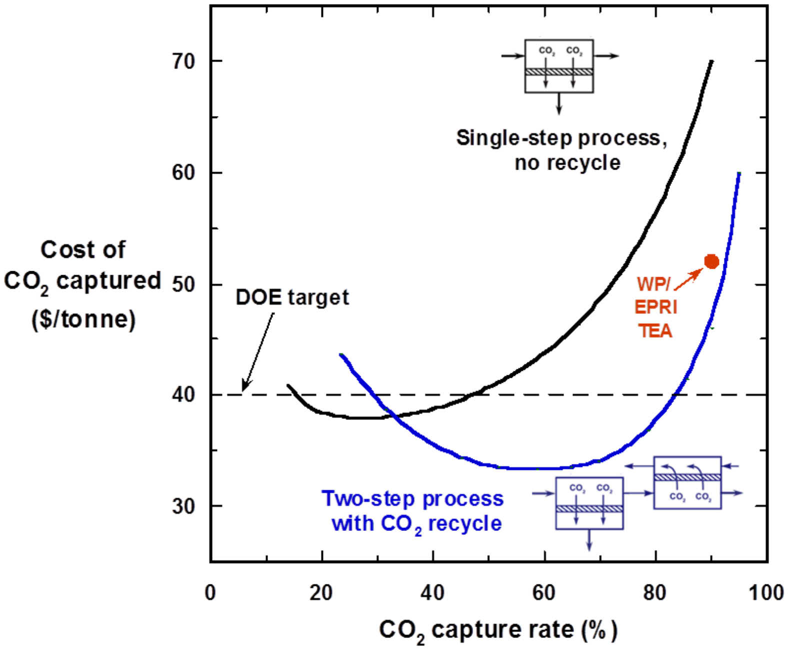


Figure ES 9. Cost of capture as a function of capture rate for two variations of the MTR membrane process. The blue curve is calculated for the baseline two-step design with selective CO_2 recycle. The black curve is for a one-step, no recycle design. Also shown in the plot is the data point calculated by WP/EPRI at 90% capture.

Industrial CO_2 Capture Field Test

While the power industry produces the majority of non-transportation man-made CO_2 emissions, there are large industrial CO_2 emission sources. For example, steel production, cement manufacture, and chemical refineries each generate between 5% and 10% of worldwide CO_2 emissions from stationary sources. Unlike the power industry, there is no renewable or decarbonized means of producing these valuable industrial materials. As a result, CO_2 capture options will be needed to reduce emissions from industrial sectors. Membranes are an attractive industrial capture approach for several reasons. For example, membranes do not use steam (in contrast to solvent capture processes), which may not be as readily available in industrial settings as it is in a power facility.

Early in this project, MTR evaluated membrane system designs and estimated costs of CO_2 capture from a refinery hydrogen production process. High CO_2/H_2 selectivity Polaris membranes were found to offer capture costs of $20/tonne when combined with a conventional pressure swing absorption hydrogen purification system.

Based on the potential of Polaris membranes to be used for industrial CO_2 capture, a search was initiated to identify a host site for conducting a small validation field test. This effort led to identification of an opportunity to test at a municipal waste-to-biofuels facility operated by

Enerkem near Edmonton, Canada. Design calculations showed that two types of MTR membranes could be used for CO_2 separation and process efficiency improvements:

- a CO_2-selective Polaris membrane optimized for high pressure, cold operation would perform CO_2/H_2 separation on syngas leaving the bio-waste gasifier to debottleneck or replace a Rectisol separation system.
- a H_2-selective Proteus™ membrane can recover H_2 from a H_2/CO_2 mixture leaving a methanol reactor to improve the efficiency of the biofuels production process.

A bench-scale system (approximately 1 TPD) was designed and built to test commercial length modules of both types of membranes. Preparation work at the host site (the Advanced Energy Research Facility or AERF) was co-funded by Alberta Innovates-Energy and Environmental Solutions (AI-EES) and performed by MTR's collaborators, Enerkem, and the City of Edmonton.

Figure ES 10 shows a picture of the MTR industrial CO_2 capture test system during installation at the AERF in Edmonton. At the time of this report, the test system installation and shakedown operations on air had been completed. However, commissioning and operation was delayed due to issues with the plant providing syngas to the system. The planned 1,000 hour test campaign to evaluate membrane performance is currently expected to occur in summer 2016. An update to this report will be issued when the field test and data analysis is completed.

Figure ES 10. A picture of the MTR industrial CO_2 capture test system during installation at the host site in Edmonton.

Membrane Capture System Water Analysis

In addition to advantages like size and simplicity, membrane capture systems should use less water compared to conventional solvent capture systems. As part of this program, the water use by the MTR capture process was quantified. Figure ES 11 compares the raw water consumption for a no capture power plant (BBS Case 11) with a conventional amine process (BBS Case 12) and the base case MTR system both operating at 90% capture. Relative to the no capture case (7.7 gpm/MW_e), both capture systems have increased water demands. However, compared to a conventional amine system where the increase in raw water consumption is 83% (14.1 gpm/MW_e), the membrane system shows a significantly lower water demand (11.5 gpm/MW_e or a 49% increase over the no capture case). As a result, the membrane system uses 18% less water than the amine process. This relatively low water demand for the MTR process can be attributed to membranes being a pressure-driven separation rather than a temperature-driven separation, which carries additional cooling water demands.

Another finding from the water analysis is that the MTR membrane process effectively harvests water from flue gas. As a result, the flue gas vented to the stack contains 85% less water when membranes are used for CO_2 capture as compared to the conventional amine capture approach. This difference amounts to an additional ~40,000 kg/h of water recovered by the membrane capture process. Reuse of such water within the power plant may have benefits particularly in arid regions, although these potential benefits were not defined in this program.

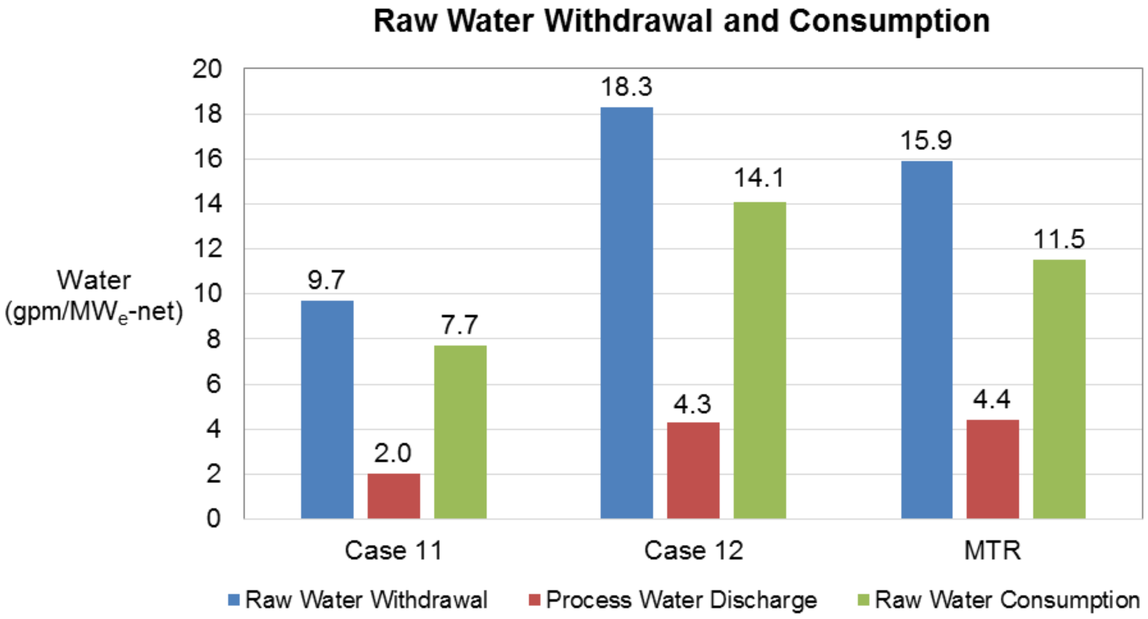


Figure ES 11. Water withdrawal and consumption for the MTR base case design at 90% capture compared to a no capture plant (BBS Case 11) and a conventional amine system (BBS Case 12) offering 90% capture.

Conclusions and Recommendations

This project resulted in the successful scale-up of the MTR membrane post-combustion capture process to the small pilot stage, including the design, build and installation of a 20 TPD capture unit, and culminating in a 1,000 hour field test of this system treating coal-fired flue gas at the NCCC. In addition to this primary accomplishment, the following were achieved:

- A Gen-1 Polaris membrane (CO_2 permeance = 1,000 gpu) accumulated over 10,000 hours of operation on flue gas using a bench-scale 1 TPD system at NCCC, providing valuable membrane lifetime information.
- A Gen-2 Polaris membrane (CO_2 permeance = 1,700 gpu) with performance exceeding project targets was scaled up to commercial roll-to-roll production and validation tested on the 1 TPD system at NCCC. An advanced Polaris membrane (CO_2 permeance = 3,000 gpu) was developed at lab scale.
- Membrane module production costs were cut by 60% to $\$200/m^2$, meeting project targets and showing a pathway to eventual commercialization cost goals ($\$50/m^2$).
- A newly-designed plate-and-frame sweep module was tested at NCCC on the 20 TPD system and demonstrated to have >4 times lower sweep-side pressure drop compared to spiral modules. This improvement will save ~10 MW_e of blower energy at full scale.
- CFD modeling and pilot boiler testing at B&W confirmed the viability of selective CO_2 recycle sweep operation. The impact of recycled CO_2 on boiler efficiency was quantified (0.75% efficiency loss) and incorporated into the project TEA.
- The project TEA shows that the capture cost at 90% capture for the MTR membrane process ($53/tonne) is competitive with the baseline solvent capture approach. A sensitivity study indicates that the membrane process shows a minimum capture cost around $35/tonne at partial capture (40-60%).
- An industrial capture host site was identified at a waste-to-biofuels facility near Edmonton, Canada. A 1 TPD test system for evaluating advanced CO_2-selective and-H_2-selective membranes was designed, built, and installed. Operation has been delayed until summer 2016 due to issues with the facility providing the feed gas slipstream.
- A water consumption analysis indicates that the MTR membrane process uses about 18% less water than the base case MEA process. In addition, the vent gas from a membrane system contains 85% less water than the flue gas vent from the MEA case, and this additional recovered water may find beneficial reuse within the plant.

In summary, these accomplishments have helped to clarify the competitiveness of the MTR membrane capture approach and advanced the technology through TRL 6.

Going forward, we recommend the following steps to minimize risk and make membrane-based CO_2 capture a viable commercial option in the near future:

- At the next stage of scale-up, it will be useful to operate the membrane system over a range of capture rates including those consistent with expected EPA regulations. Essentially all of the industrial inquiries that we receive on CO_2 capture ask about either the lowest capture cost regardless of capture rate or capture systems to meet proposed EPA regulations. Testing at these partial capture conditions will allow for optimization and refined cost estimates, which is particularly important for technologies like membranes where system design and costs can change significantly depending on CO_2

removal requirements. Moreover, although B&W have shown that selective recycle to the boiler is feasible, this approach still introduces some additional risk for the MTR process that can be removed at lower capture rates by using a single stage system.

- As described in this report, substantial progress has been made in improving membrane performance and reducing costs. However, to meet nth plant targets, additional membrane and module manufacturing improvements will be required. It will be important to continue these advanced manufacturing optimizations during the next stage of scale-up.
- In addition to membrane improvements, there is potential to reduce overall capture costs through balance of plant optimizations. For example, most membrane – as well as some sorbent – capture approaches utilize vacuum pumps to achieve CO_2 separation. These will be very large vacuum machines making up a significant fraction of the cost and energy use of the capture plant. However, because there is no near-term market or funding available, there is no organized effort to improve vacuum efficiency and costs. This is in contrast to CO_2 compression, where substantial research funding has been devoted to optimizing this compression equipment. Better vacuum pumps would be beneficial to membranes and other capture technologies.
- The potential benefits of co-capture of flue gas water along with CO_2 by the membrane process should be examined further. Recent studies have investigated the benefits of harvesting and dehydrating flue gas and using the collected water to allow a greater flexibility in siting power plants in arid locations. Examples of these programs include, U.S. DOE (through the *Innovative Water Reuse and Recovery* program), Southern Research Institute (in collaboration with EPRI through the *Water Research Center*), and the European Union (through the *CapWa* program). These organizations sponsored studies and R&D pilot plants using chilling/condensation, liquid desiccant, and membrane technologies. All of these programs found beneficial use for the harvested water.
- While much of the research focus on post-combustion CO_2 capture has been on coal-fired power generation, North America has seen a dramatic shift to natural gas-based power production in the past few years. This trend seems likely to continue, and eventually capture from natural gas power plants will be required to meet CO_2 emission reduction obligations. MTR has described a selective exhaust gas recycle (SEGR) approach for membrane-based capture from natural gas that is analogous to the coal process described in this report. This SEGR concept has been well-received in literature, but not yet tested because funding is required to get turbine manufacturers to participate in an evaluation of the process. We believe a feasibility study to flesh out this SEGR concept as either an all membrane approach or as a hybrid capture system would provide valuable information to the capture community.

1. INTRODUCTION

The main objective of this project was to demonstrate an efficient membrane process to capture CO_2 from a coal-fired flue gas slipstream at the small pilot scale (20 tonne CO_2/day [20 TPD] or approximately the CO_2 generated by 1 MW_e of coal-fired power production). The project work was conducted by Membrane Technology and Research (MTR) and our collaborators, Babcock and Wilcox (B&W), the National Carbon Capture Center (NCCC), the Electric Power Research Institute (EPRI), WorleyParsons (WP), the Illinois Sustainable Technology Center (ISTC), and Enerkem (EK) over three budget periods from 10/1/2010 to 9/30/2015. In addition to the small pilot demonstration testing at the NCCC, project work included laboratory membrane and module development at MTR, slipstream validation testing on a 1 TPD bench-scale membrane system at NCCC, boiler modeling and testing at B&W, a TEA by EPRI, WP and MTR, a case study of the membrane capture technology applied to a ~20 MW power plant in Illinois, and an industrial CO_2 capture test at an Enerkem waste-to-biofuel facility.

In sum, these activities have advanced the MTR post-combustion membrane CO_2 capture technology through TRL 6, and significantly clarified the relative potential of this membrane capture approach. This introductory chapter provides a brief background on post-combustion CO_2 capture with membranes, a history of the MTR Polaris CO_2 capture membrane and process development, and outlines the overall report structure.

1.1 Background

Over the past decade, DOE has funded a substantial research effort to identify low-cost ways to capture CO_2 from the emissions of large point sources, such as power generation facilities, to mitigate the climate impact of unabated CO_2 emissions. Coal-fired power plants have been a particular focus for CO_2 capture efforts because of the large installed base of these plants, which produce almost 40% of U.S. CO_2 emissions. In addition, the relative low cost and large domestic supply of coal suggests that this fuel will remain important to power production for the foreseeable future.[1, 2]

Currently, amine absorption is the leading candidate technology for post-combustion CO_2 capture. This first-generation (Gen-1) capture approach is a proven technology used successfully to remove CO_2 from industrial gas streams for decades. However, a number of studies have shown that amine absorption, when applied to flue gas CO_2 capture, is going to be costly and energy intensive.[2, 3] For example, estimates indicate that for 90% CO_2 removal from coal flue gas, the cost of capture for Gen-1 amines will be ~\$80/tonne CO_2. In addition, there are health and environmental concerns about handling large amounts of toxic amine solvents as well as emissions of these compounds from the capture plant. As a result, DOE is funding development of a suite of second-generation (Gen-2) technologies based on advanced solvents, sorbents, and membranes, with the ultimate goal of reducing the capture cost to \$40/tonne or less.[2, 4]

Among the Gen-2 capture technologies being developed are a number of membrane approaches.[5-8] Membrane processes offer some advantages when applied to post-combustion

CO_2 capture, including no hazardous chemical storage, handling or emissions issues, no modifications to the existing power plant steam cycle (because they use only electricity rather than steam), simple passive operation, tolerance to high SO_x and NO_x content, recovery of flue gas water, and membranes are particularly cost-effective at partial capture. The main challenge for post-combustion capture membranes is the low partial pressure of CO_2 in flue gas, which results in very large membrane area being required because of the small driving force for separation. Some years ago, working with DOE, MTR made two key innovations to address this problem:

1. New membranes, called Polaris, that have ten times the CO_2 permeance of conventional gas separation membranes were developed. A tenfold increase in permeance leads to a tenfold decrease in the required membrane area, and reduces the capital cost and footprint of the capture system substantially.
2. A membrane selective recycle process was developed. This patented process uses combustion air as a sweep stream to generate driving force for transmembrane CO_2 transport.[9] The separated CO_2 is recycled to the boiler with air. This design increases the concentration of CO_2 in flue gas, which reduces the energy and capital required for subsequent capture.

Over the past 8 years, MTR has worked with DOE to develop these innovations into a cost-effective CO_2 capture process. Figure 1.1 summarizes the timeline of this development effort. An initial project starting in 2007 (DE-NT43085) first proposed and then examined in modeling studies the feasibility of a selective recycle membrane process. During this project, the Polaris membrane configuration was also first conceived and tested in the lab. This work led to a follow-on project (DE-NT0005312) that included the first test of membrane modules with coal-fired flue gas at the Arizona Public Services (APS) Cholla plant in 2010. The APS Cholla test utilized a 50 kW_e or 1 tonne CO_2/day (1 TPD) membrane unit that housed 8-inch diameter spiral wound Polaris modules. This test demonstrated that Polaris membranes were robust enough to survive months of coal flue gas treatment without showing degradation, and it was the first field validation of air sweep module performance. After completion of the Cholla project, the current program (DE-FE0005795) was initiated with a primary goal of scaling up the MTR capture approach to a 1 MW_e or 20 TPD small pilot test system. As part of this project, the 1 TPD Cholla unit was refurbished and moved to the NCCC so that membrane lifetime data while treating coal flue gas could be collected. By the end of this project, the 1 TPD system had accumulated over 10,000 hours of operation treating flue gas at NCCC (described in more detail in Chapter 3). In 2014 and 2015, the 20 TPD small pilot system was installed and operated at NCCC (detailed results are discussed in Chapter 4). This system operated successfully with flue gas for over 1,000 hours demonstrating advanced bundled spiral wound modules for the vacuum capture step, as well as low-pressure-drop, plate-and-frame modules for the air sweep step (these sweep modules were developed separately under DE-FE0007553). Also as part of the 20 TPD program, B&W conducted modeling and testing of coal-fired boiler operation with CO_2-laden air to mimic the MTR selective recycle process. Based on the encouraging findings from these studies, a new project (DE-FE0026414 started in July 2015) will utilize the 20 TPD small pilot system – operating in an integrated fashion with a B&W research boiler – to test the complete MTR capture process including CO_2 recycle to the boiler. These activities over the past 8 years have brought the MTR membrane technology from concept (TRL 2/3) through small pilot validation (TRL 6).

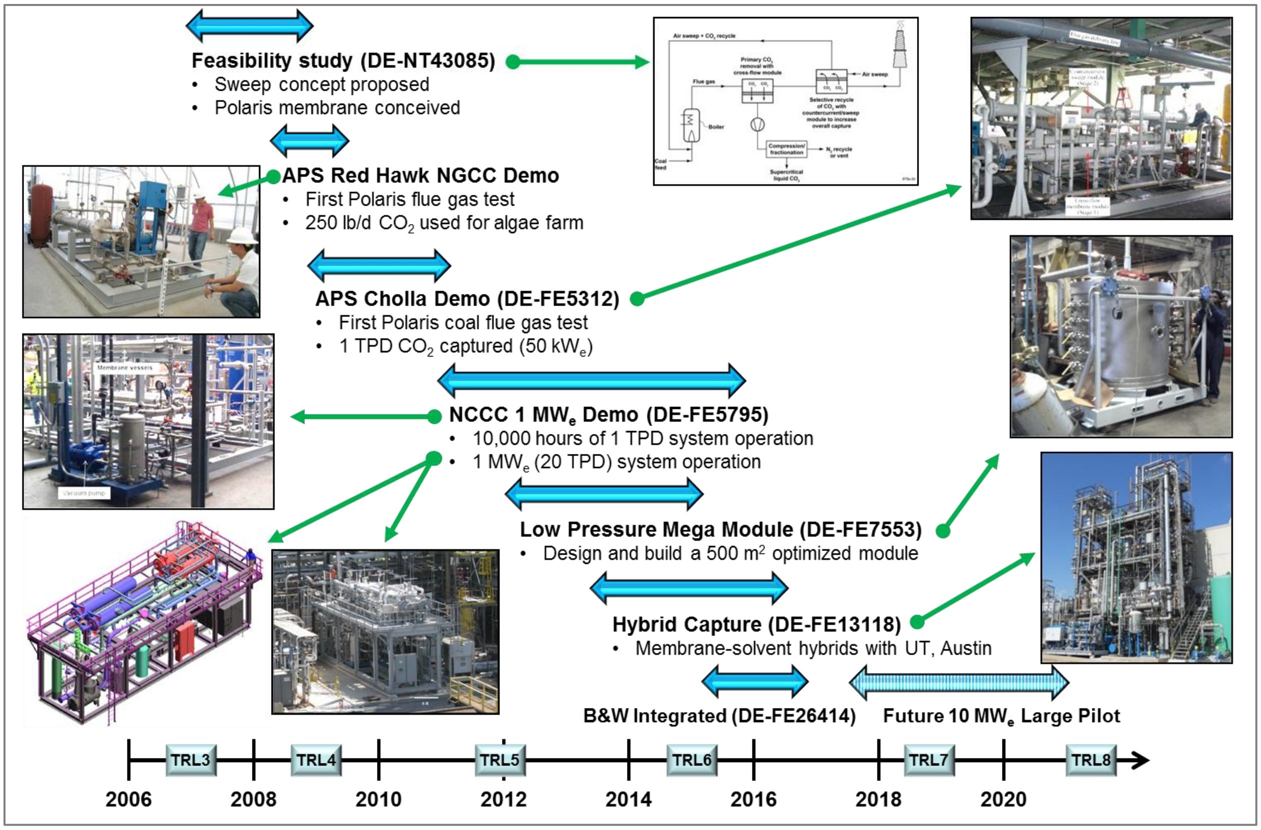


Figure 1.1. MTR Polaris membrane CO_2 capture process development timeline.

1.2 Membrane Fundamentals

Polymer membranes separate the components of a gas or vapor mixture because the components permeate the membrane at different rates. The permeability, P [cm^3(STP)·cm/cm^2·s·cmHg], of a polymer membrane material for a gas is defined as the rate at which that gas moves through a standard thickness (1 cm) of the material under a standard pressure driving force (1 cmHg). A related parameter used more frequently in the membrane industry is gas permeance, where permeance = permeability/thickness. The permeance is frequently expressed in gas permeance units (gpu), where 1 gpu = 10^{-6} cm^3(STP)/(cm^2 s cmHg). The higher the membrane permeance, the more gas that can be treated by a given membrane area. Permeance can be increased by either increasing intrinsic permeability through changes to the membrane chemistry or by reducing the thickness of the membrane selective layer.

The separating ability of a membrane is determined by the selectivity, $\alpha_{1/2}$, defined as the ratio of the gas permeabilities, P_1/P_2, or permeances. Selectivity can also be expressed as

$$\alpha_{1/2} = \frac{P_1}{P_2} = \left(\frac{D_1}{D_2}\right) \times \left(\frac{S_1}{S_2}\right) \tag{1}$$

where D is the diffusion coefficient of the gas in the membrane (a measure of the gas mobility), and S is the sorption coefficient, which links the concentration of the gas in the membrane to the pressure in the adjacent gas. In glassy polymers, the dominant contribution to selectivity is the ratio of the diffusion coefficients, D_1/D_2, which depends on the ratio of the molecular sizes. In rubbery polymers, the dominant contribution is from the ratio of the sorption coefficients, S_1/S_2, which is proportional to the ratio of the permeant condensabilities. CO_2 is both smaller than nitrogen and much more condensable, so membranes are always selective for CO_2 over N_2 to varying degrees. All membranes used commercially for industrial gas separations, including the Polaris membranes developed by MTR, operate by the solution-diffusion mechanism described above.

Figure 1.2 illustrates the structure of a typical thin-film composite membrane, such as that used for the MTR Polaris membrane. A microporous support material, with low resistance to gas permeation, provides mechanical strength for the membrane. The microporous support is often coated with a highly permeable gutter layer, which improves the compatibility between the support and selective layer, as well as conducting the permeating gas to the support membrane pores. The gutter layer is then coated with a selective layer composed of polymers with desirable properties for CO_2/N_2 separation. The overall separation performance of such composite membranes largely depends on the properties of the selective layer, including its permeability, thickness, and integrity. However, if the selective layer is very permeable, the membrane support layers can begin to impart significant resistance to transport and adversely affect overall membrane performance. As discussed later in Chapter 2, this issue can become important when developing very permeable advanced flue gas membranes.

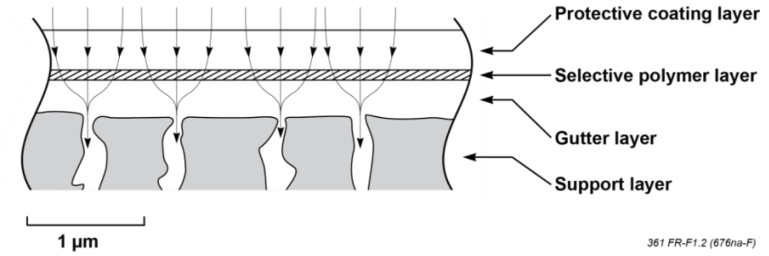


Figure 1.2. Schematic drawing of the structure of a thin-film composite membrane.

1.3 Polaris Membrane Development

The Polaris membrane has a multilayer composite structure with a selective layer based on polar polymers that are extremely permeable to CO_2 and other polar species. Figure 1.3 shows a trade-off plot of CO_2/N_2 selectivity versus CO_2 permeance for several versions of the Polaris membrane. The Gen-1 Polaris membrane set the standard against which all post-combustion capture membranes are compared. With an average CO_2 permeance of 1,000 gpu and a CO_2/N_2 selectivity of 50, Polaris was a step-change improvement over typical commercial CO_2-selective membranes used for natural gas treatment (which offer a CO_2 permeance of around 100 gpu combined with a CO_2/N_2 selectivity of 30). During this project, the Gen-1 Polaris membrane was scaled up to commercial production quantities and used in the 20 TPD small pilot system. In addition to being utilized for these coal flue gas slipstream tests at NCCC, this version of the Polaris membrane has also been employed in commercial natural gas and refinery membrane applications.

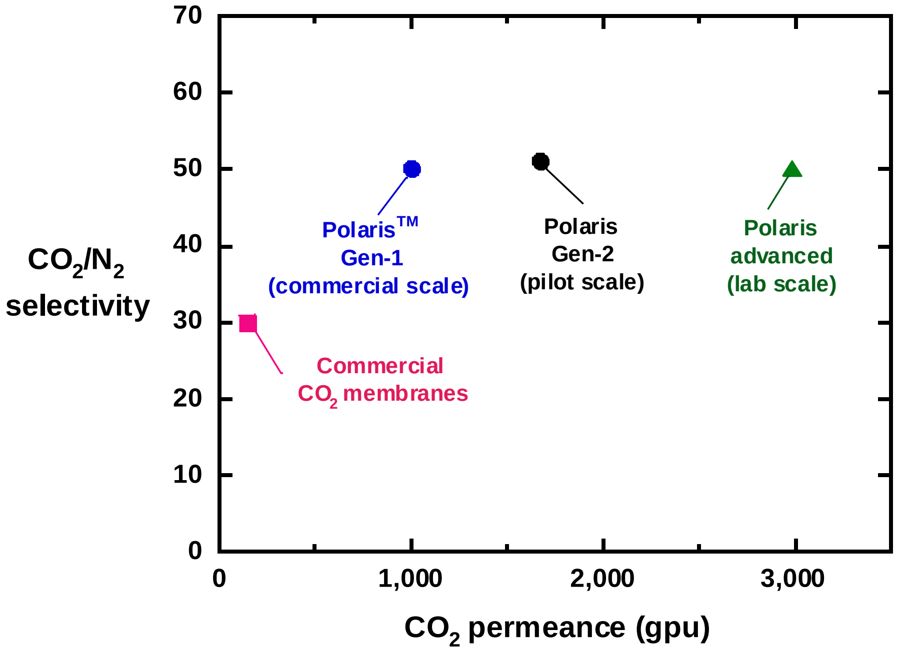


Figure 1.3. A CO_2/N_2 trade-off plot showing data for several generations of MTR Polaris, compared with the properties of a good commercial natural gas membrane. Data are pure-gas values at room temperature.

Figure 1.3 also shows some of the more recent improvements in the performance of Polaris membranes accomplished during this project (details of this membrane development and testing are described in Chapters 2 and 3). A Gen-2 version of the Polaris membrane was scaled up to pilot production. This membrane offers a CO_2 permeance of about 1,700 gpu with selectivity values similar to the base case Polaris. These Gen-2 membranes have been made on commercial casting and coating equipment, fabricated into full-sized modules, and validated in field testing at NCCC. Recently, advanced Polaris membranes with a CO_2 permeance of 3,000 gpu have been produced at the lab scale. These improvements are important because the size and capital cost of the membrane skids scales almost linearly with membrane CO_2 permeance. Thus, the advanced Polaris membranes would yield a system with one-third the number of membrane vessels as the Gen-1 membranes. The impact of these membrane performance improvements on the cost of electricity for a plant using a membrane capture system is described later in Section 1.6.

1.4 Process Design Considerations

In addition to a membrane with good separation performance, an energy-efficient and affordable process design is required to make membranes competitive for post-combustion CO_2 capture. Membrane process design studies from previous MTR/DOE projects have produced the following general conclusions about using membranes for post-combustion capture:[5]

- To capture CO_2 from flue gas, a membrane process needs partial pressure driving force. This driving force can be generated by either (a) compression on the feed side or (b) a vacuum on the permeate side of the membrane. Calculations show that the energy required is considerably lower for a vacuum process because the vacuum only has to pump the flue gas that permeates the membrane (about 10% of the total flue gas, and largely CO_2), whereas a feed compressor has to pressurize all of the flue gas (CO_2 plus the bulk N_2). A vacuum process uses less energy but requires a much larger membrane area, because the CO_2 partial pressure difference across the membrane is small. Consequently, an energy-efficient vacuum-driven process requires very permeable membranes.
- In addition to large membrane area or power requirements, single-stage membrane designs are unable to produce high-purity CO_2 *combined* with high CO_2 capture rates. In fact, a single-stage membrane process alone cannot produce high-purity CO_2 in the permeate with 90% CO_2 capture, regardless of the membrane selectivity. This is because the system performance is limited by the pressure ratio across the membrane.

In practical separation applications, the pressure ratio – that is, the ratio of the feed pressure to the permeate pressure – is usually between 5 and 15. Higher pressure ratios can be achieved by using larger compressors on the feed gas or larger vacuum pumps on the permeate; however, for flue gas treatment, energy and equipment costs limit the maximum affordable pressure ratio to about 10. Under these conditions, high membrane permeance is more important than high selectivity. In typical membrane processes, the point of diminishing returns is reached when the selectivity is about three to five times the pressure ratio.[10] For flue gas CO_2 capture, this corresponds to a CO_2/N_2 selectivity of 30-50. Higher selectivity, at the expense of membrane permeance, will only increase the required membrane area while producing little improvement in product purity.

Because of this pressure ratio constraint, and the desire to achieve a high CO_2 capture rate and high purity (90% capture at >95% purity), flue gas treatment with membranes requires a multi-step/stage design and/or hybrids with other separation technologies (cryogenics, absorption, etc.). The MTR solution to this capture challenge is the selective recycle process design shown in Figure 1.4. This process uses a combination of a small amount of feed compression and permeate vacuum in a first membrane step to efficiently generate a pressure ratio that will lead to capture of about 50% of the inlet flue gas CO_2. The partially-treated flue gas leaving this primary CO_2 removal unit is then sent to a second membrane step that utilizes a sweep gas of combustion air to selectively recycle CO_2 to the boiler and drive the overall CO_2 recovery up to 90%.

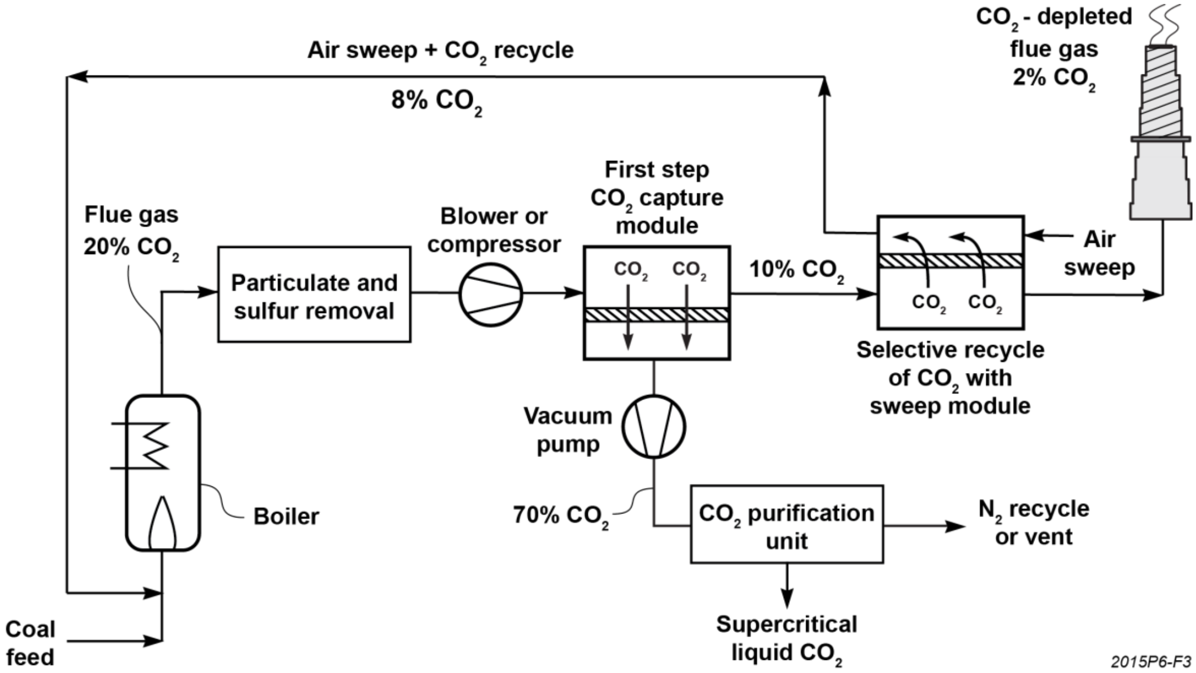


Figure 1.4. Simplified diagram of the MTR selective recycle CO_2 capture process at a coal-fired power plant.

The Figure 1.4 design has a number of features that optimize system performance:

- Because it is a two-step membrane design, all of the flue gas CO_2 does not have to be removed in a single membrane step. This allows the first-step membrane to operate efficiently at low stage-cut with a relatively high partial pressure of CO_2 on the feed side.
- The second membrane step performs the difficult task of removing CO_2 to very low levels (i.e., to reach 90% capture). This step uses an air sweep stream to maintain separation driving force by keeping a relatively low partial pressure of CO_2 on the permeate side. Because the air stream is already being blown into the boiler as the oxidant for combustion, this sweep gas provides an essentially free separation (i.e., no compressors or vacuum pumps are used in this step).
- The concentration of CO_2 in the flue gas leaving the boiler is increased (for example, from 12% to 20% CO_2) because CO_2 is recycled to the boiler with the air sweep stream. This enrichment makes CO_2 capture in the first membrane step easier due to the higher CO_2 partial pressure.
- Finally, the CO_2-enriched permeate leaving the vacuum pump is sent to a compression and purification unit (CPU) where the concentrated CO_2 can be readily liquefied using refrigeration to produce a high-purity (>99%) product ready for utilization or storage.

In particular, the use of combustion air sweep to selectively recycle CO_2 to the boiler is an effective way to reduce the minimum energy required by the capture step. As described later in Section 1.6, techno-economic studies suggest this membrane process is a competitive post-combustion capture option.

1.5 Membrane Module Design

One of the key issues for a membrane post-combustion capture system is how to balance the desire for a small system footprint with the need to process large volumetric flows and minimize parasitic pressure drops. The pressure drop issue is particularly important because for a full-scale (550 MW_e power plant) membrane capture system, each 1 psi of pressure drop through the membrane unit amounts to 2-3 MW_e of required blower energy. Based on modeling, we have estimated a 1.5 psi pressure drop through each of the membrane steps shown in Figure 1.4 (i.e., this is the pressure drop from feed to residue in each of the two membrane steps, as well as the sweep-in to sweep-out pressure drop on the air sweep step). The ability to reach this pressure drop target on a full-scale capture unit while maintaining a compact system size and good membrane performance depends on the membrane module design.

In previous CO_2 capture work with DOE, we focused on using spiral-wound modules adapted from those used for high-pressure CO_2 separations, such as natural gas treatment. Figure 1.5 shows a drawing of a conventional spiral-wound module used in the CO_2 capture step of the Figure 1.4 design, as well a modified spiral used for the selective recycle sweep step. Use of this existing module technology for flue gas CO_2 capture allowed for rapid development of the process, evaluation of the chemical and mechanical stability of the membranes, and optimization of membrane permeation properties. However, these early modules were not optimized for low-pressure flue gas CO_2 capture, particularly for the selective recycle step where a sweep gas is used. As a result, with the aid of computational fluid dynamics, the module configuration for low-pressure flue gas treatment was redesigned in this project.

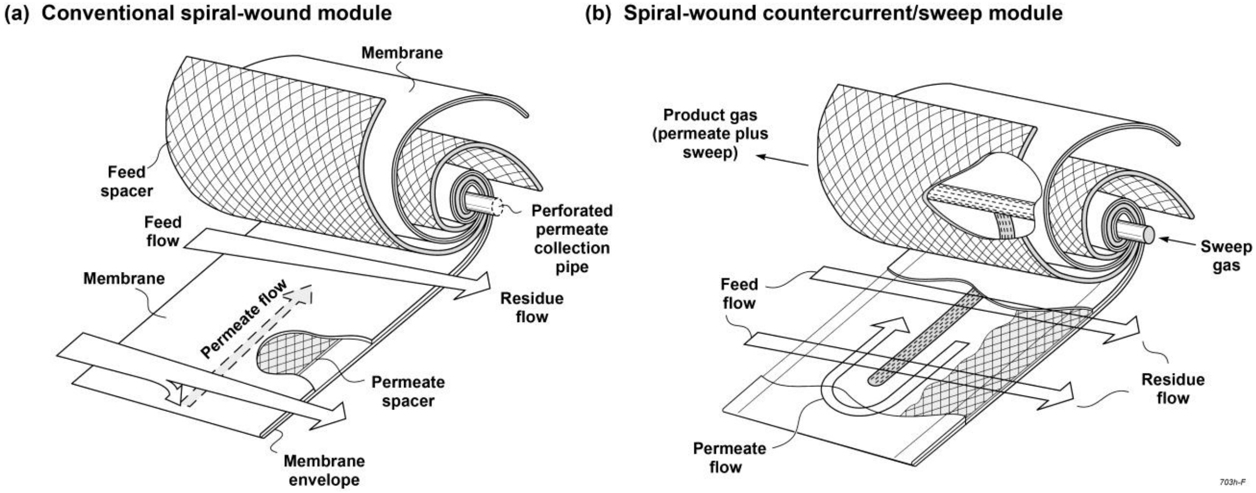
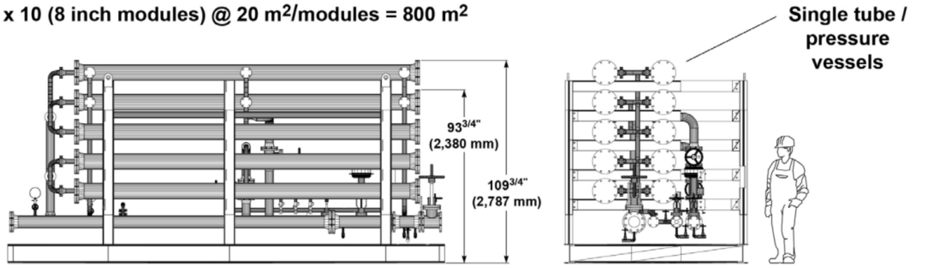


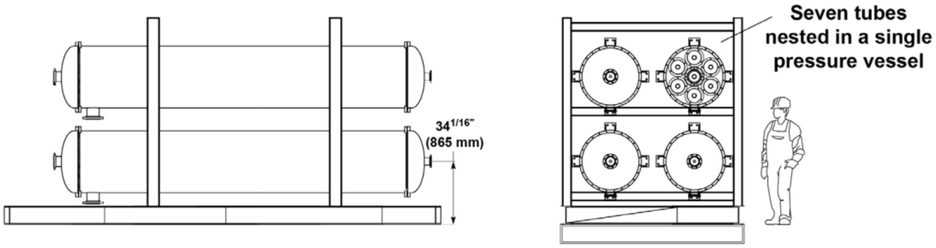
Figure 1.5. Feed and permeate flow patterns in (a) a conventional spiral-wound module and (b) in a spiral modified for sweep operation.

Figure 1.6(a) shows a drawing of a typical membrane module skid used for high-pressure CO_2 removal from natural gas. In this application, 8-inch (20 cm)-diameter, 40-inch (102 cm)-long spiral-wound modules, each containing about 20 m^2 of membrane area are housed in a large number of individual vessels. Each of the pressure vessels in the Figure 1.6(a) example houses 4 modules in series. This configuration works well when the feed-to-residue pressure drop is relatively unimportant, and the individual vessel diameter should be kept small (8 inches) to minimize the wall thickness required to safely handle the high feed pressure. However, a drawback of the Figure 1.6(a) design is that the skid holds a relatively small amount of membrane area (low packing density). When this fact is combined with the large number of valves and instrumentation required for the many pressure vessels, the installed membrane skid cost on a \$/m^2 basis is high. This was the type of skid used for the 1 TPD test system at NCCC.

(a) 4 x 10 (8 inch modules) @ 20 m^2/modules = 800 m^2



(b) 4 x 4 x 7 (8 inch modules) @ 20 m^2/modules = 2,240 m^2



(c) 4 x (1 x 1 x 5 m) modules @ 2,500 m^2/modules = 10,000 m^2

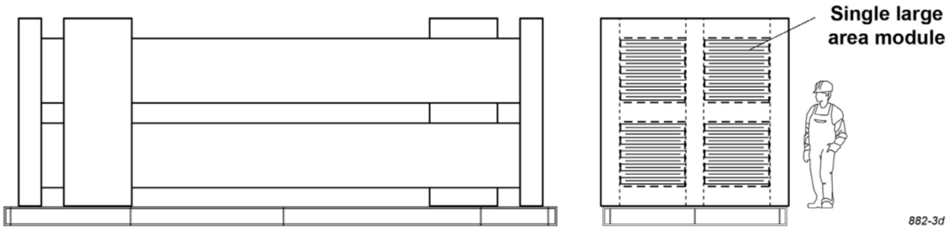


Figure 1.6. Membrane module designs.

Because the MTR CO_2 capture membrane process is a low-pressure design, the restrictions on module housing size and geometry can be relaxed compared to those needed for high-pressure membrane gas separations. Figure 1.6(b) shows a drawing of a multi-insert skid design that was used in this project for scale-up to the 20 TPD small pilot system at NCCC. This nested module design has a number of benefits; it provides higher skid packing density and a reduced number of interconnections and valves (reduced cost), as well as being more amenable to arranging membrane area in parallel rather than in series, which reduces pressure drop. Details of the design and operation of this nested module at NCCC are described in Chapter 4.

Figure 1.6(c) shows an alternative low-pressure module design. This rectangular or plate-and-frame configuration can achieve relatively high skid packing density. In addition, the regular geometry of this module is amenable to automated fabrication methods, which will reduce cost. Perhaps the most important feature of the plate-and-frame configuration is the ability for fine control of the flow path on both the feed and sweep sides of the membrane, which can be used to minimize pressure drop in the selective recycle sweep membrane step. In a separate DOE project (DE-FE0007553), we designed a large plate-and-frame module of the type shown in Figure 1.6(c) optimized for low-pressure sweep operation.

During this project, both the nested spiral wound module for the vacuum permeate step and the plate-and-frame module for the air sweep step were demonstrated for the first time on the 20 TPD small pilot. We believe these optimized low-pressure module configurations will be used for future large capture systems. Details of their impact on system cost and performance are provided later in this report.

1.6 Techno-Economic Studies

Over the years, a number of techno-economic analyses (TEAs) of the MTR membrane post-combustion capture process have been conducted. For example, a recent DOE TEA report on future technologies for post-combustion carbon capture compares the MTR membrane approach favorably with various amine processes.[4] This study shows a membrane system using advanced Polaris membranes (CO_2 permeance of 3,500 gpu) and advanced compression equipment (93% efficiency) capturing 90% of the CO_2 from an ultra-supercritical coal plant while approaching the DOE target of a 35% increase in cost of electricity (COE) and a cost of CO_2 avoided of <$40/tonne.[4]

During the first budget period of this project, MTR completed a TEA that included an examination of the impact of membrane performance on the COE for a membrane system operating at 90% capture. Figure 1.7 shows the results of this analysis where the methodology described in the Bituminous Baselines Study [3] was followed. The change in the COE when the membrane capture system is employed is quite sensitive to the membrane CO_2 permeance, particularly at lower permeance values. For the Gen-1 Polaris membrane (1,000 gpu), the increase in COE is about 70%. This value is slightly better than the DOE base case amine process (MEA), similar or slightly worse than some advanced amines, and far from the DOE target of 35%. For Gen-2 Polaris membranes (1,700 gpu), the change in COE is reduced to 55-60%, a significant improvement. For advanced Polaris membranes (~3,000 gpu), the increase in COE is approaching 40%, just above the DOE target. At still higher membrane permeances, there are diminishing returns, as the membrane starts to comprise a small portion of the overall system cost.

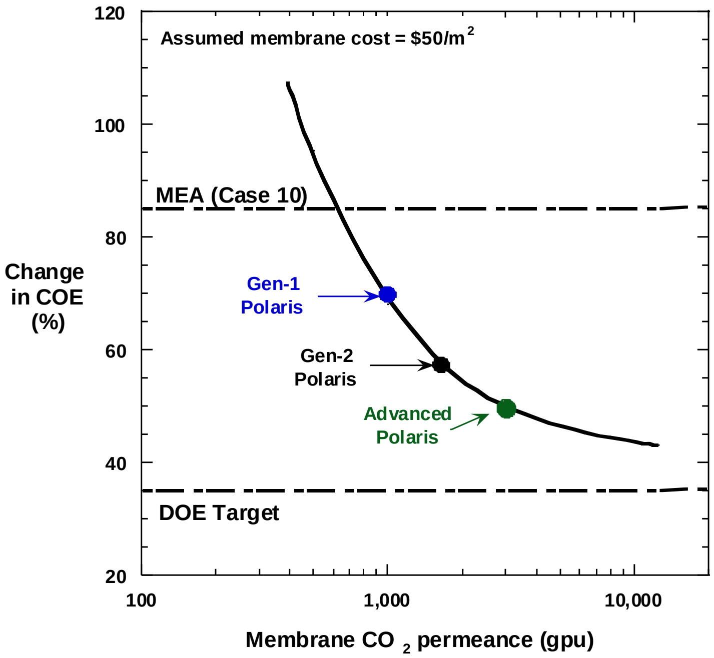


Figure 1.7. Effect of membrane CO_2 permeance on the increase in the COE at 90% CO_2 capture from a pulverized coal power plant using the MTR capture process. The membrane CO_2/N_2 selectivity is 50 and the installed membrane skid cost is $\$50/m^2$.

After installation of the 20 TPD small pilot unit at NCCC, EPRI and WP conducted an updated TEA of the MTR CO_2 capture process. The results of this analysis are provided in Chapter 6.

1.7 Report Objectives and Organization

The purpose of this report is to document work conducted by MTR and our subcontractors to better understand the potential of membrane technology to be used for post-combustion CO_2 capture. This work involved laboratory membrane/module development at MTR, slipstream field testing of 1 and 20 TPD membrane systems at NCCC, boiler modeling and testing at B&W, a systems/economic analysis by EPRI, WP and MTR, a case study by ISTC of the membrane capture technology applied to a ~20 MW power plant in Illinois, and an industrial CO_2 capture test at an Enerkem waste-to-biofuel facility. Each of these topics is discussed in this report with the exception of the ISTC case study, which was summarized in a previous topical report. The remainder of this report is organized in the following manner:

- Chapter 2 describes membrane and module development at MTR;
- Chapter 3 discusses validation testing of the 1 TPD system at NCCC;
- Chapter 4 reviews the design, fabrication and operation of the 20 TPD small pilot system;
- Chapter 5 summarizes modeling and testing work to examine the impact of CO_2 recycle on boiler performance conducted by B&W;
- Chapter 6 provides a summary of the WP/EPRI/MTR TEA of the MTR capture process;

- Chapter 7 reviews the design and installation of an industrial CO_2 capture membrane system at the Advanced Energy and Research Facility in Edmonton;
- Chapter 8 summarizes an EPRI analysis of water use in the MTR capture process;
- Chapter 9 provides an Environmental Health and Safety evaluation of the MTR capture process;
- Chapter 10 provides a review of progress on project milestones, and
- Chapter 11 summarizes our conclusions, lessons learned, and recommendations for future work.

There are two appendices included at the end of the report: The first is the full comparative TEA prepared by EPRI and WP, which is summarized in the Chapter 6 discussion. The second appendix is the complete EPRI report on water management within the MTR capture process (highlights of this report are discussed in Chapter 8).

2. MEMBRANE AND MODULE PERFORMANCE IMPROVEMENT AND COST REDUCTIONS

2.1 Membrane Permeance Improvements

Power plant flue gas has a low partial pressure of CO_2 and enormous volumetric flow rates. Even using the cost-effective process design shown in Figure 1.4, calculations show that membranes must have a minimum CO_2 permeance of about 1,000 gpu (where 1 gpu = 10^{-6} cm^3 (STP)/ $cm^2 \cdot s \cdot cmHg$) and CO_2/N_2 selectivity of greater than 30 to make CO_2 capture with membranes economically feasible. Typical commercial CO_2-selective membranes used for treating natural gas have a CO_2 permeance of about 100 gpu. If membranes are to be competitive for post-combustion CO_2 capture, there is a clear need for custom membrane development. Recognizing this challenge, in previous DOE-funded work, MTR developed Polaris^{TM} membranes with CO_2 permeances of 1,000 gpu and CO_2/N_2 selectivities of 50. While this performance set the standard for post-combustion membranes, there is a compelling reason to strive for higher membrane CO_2 permeance: doubling the CO_2 permeance will roughly halve the required membrane area, and thus reduce the capital cost and footprint of a membrane CO_2 capture system. The impact of these membrane improvements on the cost of electricity was shown in Figure 1.7. With this benefit in mind, we set a goal at the start of this project to achieve at least a 50% improvement in CO_2 permeance without a loss of selectivity compared to the base line or Gen-1 Polaris membranes.

Figure 2.1 summarizes the membrane performance improvements that were made in this project. The data are presented in the form of a trade-off plot, where CO_2/N_2 selectivity is plotted against CO_2 permeance. Over time, we have steadily improved the performance of Polaris membranes, particularly by increasing CO_2 permeance (data points move to the right on the figure). By the end of the project, advanced membranes with 3 times the CO_2 permeance of the Gen-1 Polaris membrane were being produced in the laboratory (3,000 gpu versus 1,000 gpu).

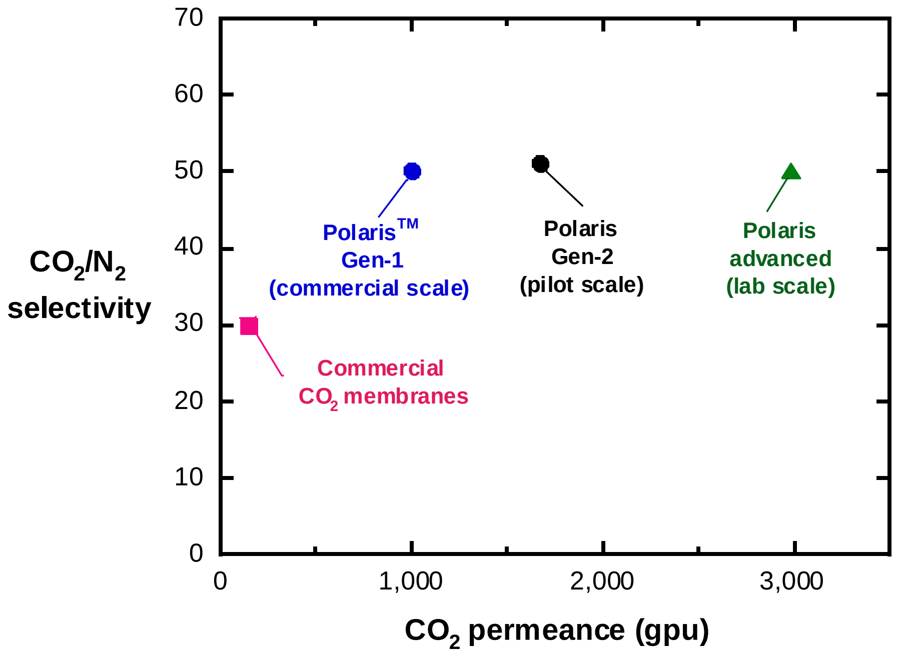


Figure 2.1. A CO_2/N_2 trade-off plot showing data for several generations of MTR Polaris, compared with the properties of a good commercial natural gas membrane. Data are pure-gas values at room temperature.

As mentioned in Chapter 1, Gen-1 Polaris membranes were used to produce the modules that were tested on the 20 TPD small pilot system. This Gen-1 membrane was fabricated on MTR's commercial roll-to-roll production equipment. An average Polaris roll is about 200 m long and 1 m wide. Over 20 production runs of this scale were conducted to make Gen-1 membrane for the 1 and 20 TPD systems. The fabrication experience gained during this project allowed the Gen-1 Polaris membrane to transition from a research development to a full commercial product that MTR now offers for various natural gas and refinery applications. Also during this project, a Gen-2 Polaris membrane with 70% higher CO_2 permeance and similar selectivity to Gen-1 was scaled up from laboratory samples to pilot-scale roll production. Currently, Gen-2 is produced on the same commercial roll-to-roll equipment as the Gen-1 membrane. Field validation of the Gen-2 membrane on the 1 TPD system at NCCC is described in Chapter 3. Finally, an advanced Polaris membrane (future Gen-3) was produced and tested on a lab scale. This membrane shows a CO_2 permeance of 3,000 gpu with similar selectivity as earlier Polaris membranes. Although resources precluded scale-up and field testing of this membrane during the project, we believe future work will allow this membrane to further reduce the costs of membrane-based CO_2 capture by the time the technology is ready for full-scale deployment.

As described in the Introduction, in theory it is possible to make membranes with even higher permeances than the advanced Polaris membranes. For example, considering that polymer materials with CO_2 permeabilities of more than 1,000 Barrer are known in the literature, and that MTR can reliably make selective layer coatings of 0.1 μm thickness, it has been suggested that 10,000 gpu (= 1,000 Barrer/0.1 microns) membranes should be a goal. However, we don't believe this is a productive use of resources for two reasons:

1. As shown in Figure 1.7 (and discussed further in Chapter 6), there are diminishing economic returns for higher performance membranes, particularly when the CO_2 permeance is greater than 3,000 gpu. At these higher membrane permeances, the membrane skid cost starts to become a relatively small portion of the overall capture plant cost, so further reductions in the membrane skid size (through permeance increases) have a relatively small influence on the capture cost.

2. Attaining very high membrane permeances in practice is challenging. The standard support layers of a composite membrane (see Figure 1.2) have been designed for gas separation membranes that are much less permeable than Polaris. For these low permeance membranes, relative to the selective layer, the support layers contribute negligible transport resistance. However, at Gen-1 Polaris performance levels (CO_2 permeance 1,000 gpu), we are approaching the point where the resistance of the support layers of the membrane begins to adversely influence the overall membrane performance.

Because of this transport limitation, and the diminishing economic returns for higher performance membranes, we believe advanced membranes with CO_2 permeance in the 3,000 gpu range is a practical long-term development target. Much of our effort in this project was directed toward modifying the support layers of the Polaris membrane so that membranes with permeances higher than 1,000 gpu could be made without sacrificing selectivity.

To better understand this support resistance issue, it is helpful to look at the resistances-in-series that affect the overall performance of a composite membrane. For example, a typical composite membrane support layer can have a CO_2 permeance of 5,000 gpu combined with low CO_2/N_2 selectivity (10 or less). When the CO_2 permeance of the selective layer is 100 gpu, as it is for conventional natural gas CO_2 membranes, the resistance due to the support layer is only 2% of the selective layer resistance. As a result, the overall membrane properties (permeance and selectivity) are essentially that of the selective layer. When the selective layer permeance is increased to 1,000 gpu, as it is for Gen-1 Polaris, the support layer provides 20% of the resistance to CO_2 transport and starts to influence the overall membrane properties. For a 5,000 gpu selective layer, the resistance of the support is now equivalent to that of the selective layer. For this case, the CO_2 permeance of the overall membrane will be cut in half (to 2,500 gpu) compared to the selective layer, and the CO_2/N_2 selectivity will be dramatically reduced (to 30 if the selective layer CO_2/N_2 selectivity is 50, and the support layer selectivity is 10). This example illustrates how producing a high-performance composite membrane is not simply a matter of coating a thinner selective layer. To realize the benefits of a higher permeance selective layer, there is a need for better (lower resistance) supports to debottleneck Polaris membrane improvements.

During this project, we designed and built a custom casting machine that allowed us to optimize substrates for very permeable membranes. Figure 2.2 shows a picture of this system during installation at the MTR laboratories. This machine was used to produce more open substrates that are used in the Gen-2 and advanced Polaris membranes.

Figure 2.2. A picture of the small casting machine used to optimize substrates for Gen-2 Polaris membranes. The casting tray and quench bath are visible at the center right of the picture. The rinsing/annealing baths and membrane take-up roll are on the left side of the picture.

As an example of the type of substrate optimization conducted during the project, Figure 2.3 shows the performance of various substrates on a CO_2/N_2 tradeoff plot. Based on our resistance-in-series calculations, we determined that a substrate CO_2 permeance of >10,000 gpu was needed to take advantage of Polaris selective layer improvements and produce a high performance (Gen-2 and better) Polaris membrane. At the same time, the substrate needs to have a minimum CO_2/N_2 selectivity of around 10 because at lower values the support is so open it becomes hard to coat a defect-free selective layer. In other words, we want the substrate to be very open (porous) so that it presents little resistance to gas transport, but at the same time these pores must be small and finely distributed so that very thin selective layers can be coated on top of the substrate without yielding defects. Each of the data points in Figure 2.3 represents a different substrate developed in this project (with the year of production as a label). All of these support membranes have higher permeance than the Gen-1 Polaris substrate, but many of them have lower CO_2/N_2 selectivity, and thus were determined to be not suitable for scale-up. However, a few of the substrates showed properties within the target performance window. These are the substrates that were used for Gen-2 and advanced Polaris development shown in Figure 2.1.

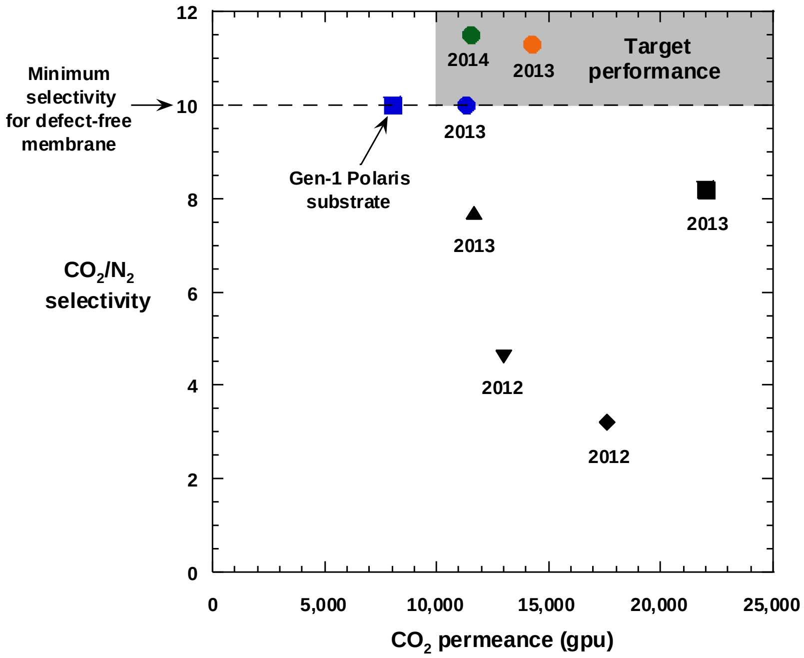


Figure 2.3. A plot showing CO_2/N_2 selectivity as a function of CO_2 permeance for Polaris membrane substrates developed in this project.

2.2 Module Performance Improvements

In addition to membrane performance improvements, work was conducted during this project to optimize modules for low-pressure drop operation. As described earlier, the energy cost to overcome the pressure drop of gas flowing through modules can be significant for low-pressure flue gas treatment. This is particularly true for the air sweep step of the MTR process. During previous bench-scale field testing of Polaris spiral-wound sweep modules, MTR found that the pressure drop on the sweep side of these modules was unacceptably high. Analysis showed that this high pressure drop occurred because sweep air flowing through the module had to make numerous turns over a relatively long flow path [as shown in Figure 1.5(b)].

During this project, a number of modifications were made to the spiral-wound sweep module design to reduce pressure drop. Figure 2.4 compares sweep-side pressure drop for various spiral wound module designs tested under the same conditions. The data indicate that the best spiral-wound sweep modules produced in this project were able to reach a sweep-side pressure drop of just under 4 psi. This was significantly better than the early spirals tested at APS Cholla (>20 psi sweep-side pressure drop). However, under equivalent laboratory conditions, plate-and-frame modules designed under DOE project DE-FE0007553 can achieve a pressure drop of less than 0.5 psi. This value is below our target of 1.5 psi used in techno-economic calculations.

Based on these comparative results, it was decided to move forward with testing a large plate-and-frame sweep module on the 20 TPD system at NCCC. These validation tests on the 20 TPD small pilot system confirmed the advantage of plate-and-frame modules for low pressure sweep operation (detailed results are described in Chapter 4). Because of this pressure drop advantage when a sweep gas is used, the plate-and-frame module design will be the future choice for the selective recycle step of the MTR CO_2 capture process.

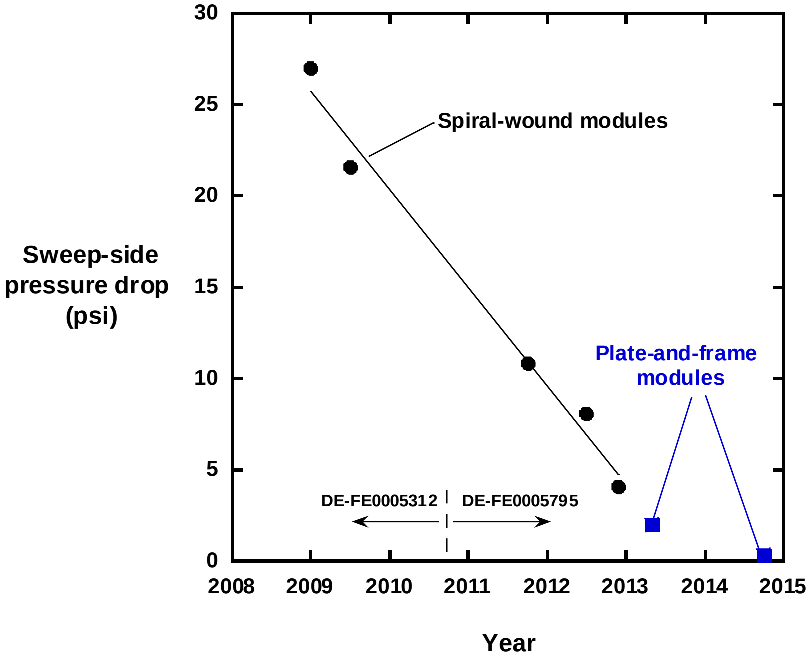


Figure 2.4. Improvements in sweep-side module pressure drop over time.

2.3 Membrane and Module Cost Reductions

One of the key issues for the competitiveness of a membrane-based CO_2 capture system is to reduce the cost of the membranes and modules, because these components make up a significant fraction of the capital cost of a large membrane system. At the start of this project, we were producing Gen-1 Polaris (1,000 gpu) on a lab scale, and packaging this membrane into high-pressure gas separation modules where the total module cost was about $500/m^2. As shown in Figure 1.7, the ultimate membrane targets to approach DOE capture cost goals in nth plant are a membrane cost of $50/m^2 and CO_2 permeance of ~3,000 gpu. The previous section highlighted our progress toward meeting the CO_2 permeance target. Here, we describe the cost reduction efforts during this project, which successfully lowered Polaris module costs to $200/m^2 (a 60% reduction).

Throughout this project, a significant effort was devoted to optimizing the process of Polaris membrane fabrication to reduce labor and materials costs. As described in the previous section on membrane improvements, a small casting machine was designed and built to optimize the Polaris support layers. We also designed and installed a small coating machine to focus on optimizing the Polaris selective layer coating. Figure 2.5 shows a picture of this coating machine installed at MTR. This small coating machine uses the same dip-coating-plus-thermal-drying technique as our large-scale commercial membrane coaters. However, the R&D coater produces short (10 to 100 feet), narrow membrane rolls (12 inches wide), so that small amounts of materials can be used while optimizing membrane formulations. Through trial and experimentation on this R&D machine, we were able to combine a number of processing steps from the original Polaris fabrication process. This had a two-fold effect of reducing the time required to make Polaris membranes (reduced labor cost) as well as a decrease in the amount of materials used in the membrane (decreased materials cost). Once these innovations were perfected in the small R&D coating machine, they were then transferred to our larger commercial coating machines. The fabrication improvements made over the course of this project were significant, as they cut the production time for Polaris membranes by more than a factor of two. The impact on the overall membrane module cost was a decrease of about 15%.

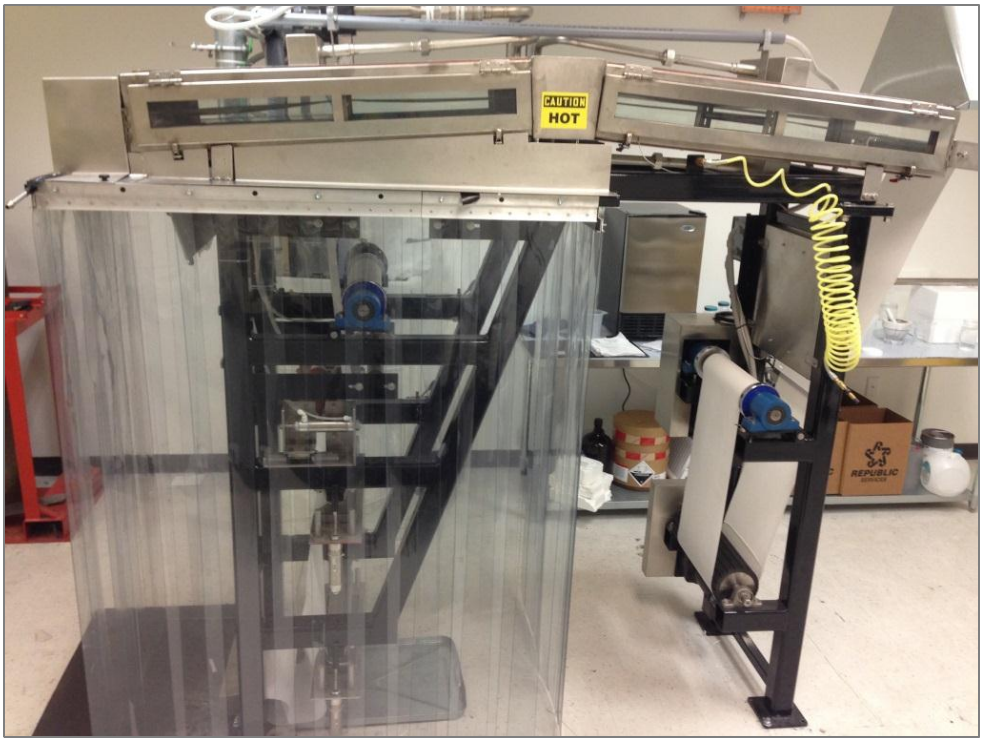


Figure 2.5. A picture of the small coating machine designed and installed in this project and used to optimize the Polaris membrane. The drying oven is on the top of the picture (marked "Hot"), the membrane take-up roll is on the right, and the coating tray is behind the plastic enclosure on the left.

Another area of cost reduction addressed in this project was to replace standard high-pressure module parts with low-cost plastic components. An example of this approach was the use of plastic module seal carriers that were designed and demonstrated on the 1 TPD system during budget period 1. These module components are used to provide a sealing surface with the module housing, and to prevent module telescoping. In typical high pressure applications, stainless steel (SS) seal carriers – costing four times as much as plastic ones – are used. Figure 2.6 shows photographs of modules using (a) the conventional SS seal carrier and (b) a low-cost plastic seal carrier. The overall effect of using the plastic seal carriers is about a 20% reduction in the fabrication costs of an 8-inch spiral-wound module. Based on the successful demonstration of these components early in the project on the 1 TPD unit, they were used on all of the modules installed on the 20 TPD small pilot system. Moreover, the cost and weight advantage of the plastic seal carriers is so dramatic that they have been adopted for use on some MTR commercial products.

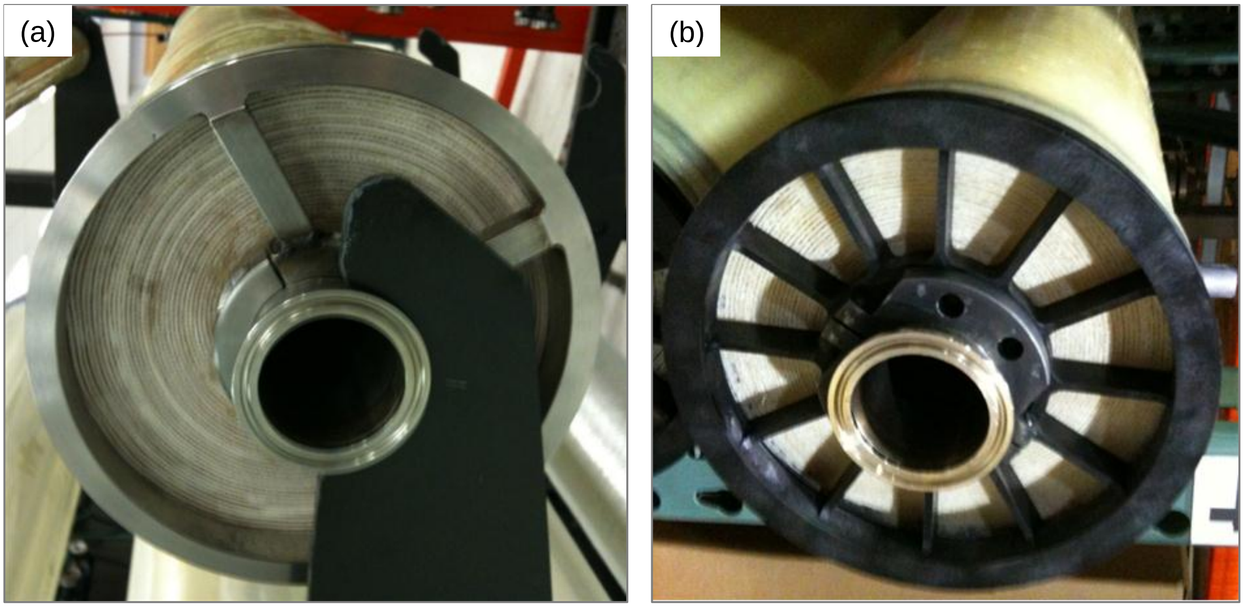


Figure 2.6. Photographs of spiral-wound module seal carriers made from (a) stainless steel (SS) and (b) plastic. The cost of the plastic seal carrier is about 25% that of the SS seal carrier.

One well-known general means of reducing manufacturing costs is to implement automation into the fabrication process. This approach has been widely deployed in the manufacture of reverse osmosis (RO) membrane modules for sea water desalination. The RO membrane industry is seen as a model for membrane post-combustion CO_2 capture because the volume of membrane and modules used in RO plants is similar to that required for a full-scale power plant CO_2 capture Polaris system. As a first step in this direction, an automated module trimming machine was designed, installed, and tested during this project. A picture of this machine during installation at MTR is shown in Figure 2.7. Module trimming, when done by hand, is a labor intensive step necessary to open module feed flow channels and achieve module size specifications. Automation of this fabrication step produces modules at a faster rate with better pressure drop performance, and results in about a 5% decrease in installed module cost. As with the plastic module components, use of the trimming machine has been adopted for many MTR commercial products.

Figure 2.7. Photograph of the membrane module trimming machine designed, commissioned, and tested during budget period 1. The machine allows modules to be fabricated at a faster rate (reduced cost) and with lower pressure drop than the previous technique.

Another cost reduction approach tested during this project was use of multi-insert module vessels. These large membrane module housings significantly reduce the number of vessels and interconnecting valves required for a given membrane area compared to conventional module vessels, such as those used on the 1 TPD system. Multi-insert module vessels were designed, built, and successfully operated on the 20 TPD system as part of this project (detailed test results are described in Chapter 4). Figure 2.8 shows one of these multi-insert vessels during fabrication when loading of the individual spiral wound modules was being tested. Our calculations show that the multi-insert module vessel reduces the installed membrane module cost of a large system by 20%.

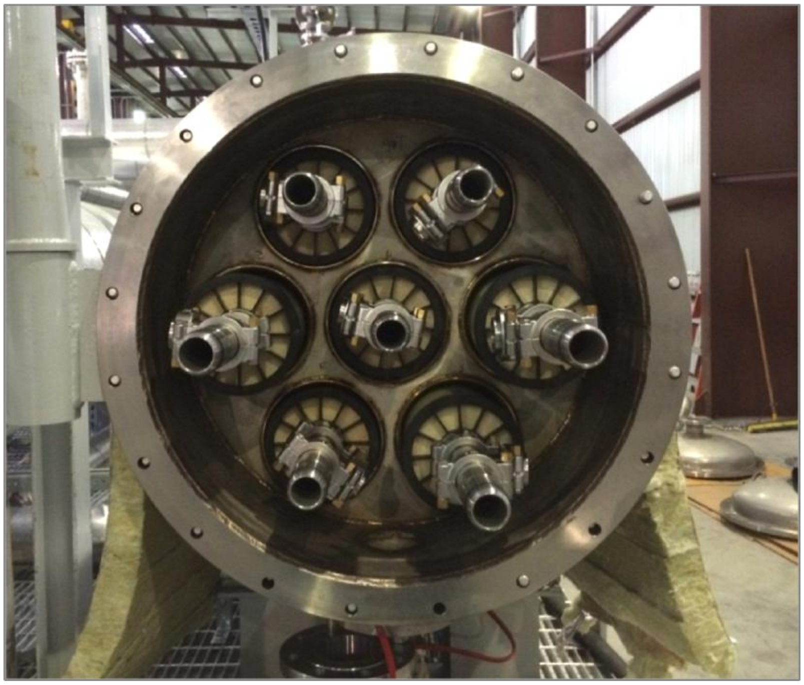


Figure 2.8. Photograph of a multi-insert pressure vessel built for the 20 TPD system during this project.

The combination of the cost reduction items described above reduced the Polaris module skid cost from \$500/m^2 to \$200/m^2. These membrane and module cost reductions were first incorporated into the Gen-1 Polaris membranes and later extended to the higher performance Gen-2 membranes. Each of these changes was validation tested with real flue gas at NCCC on the 1 TPD system (optimized membrane fabrication, low cost module components, and automated module trimming) and later on the 20 TPD system (multi-insert module vessels).

While significant progress on cost reduction has been made, there is still work to do to get from where we are today (~\$200/m^2), to where we need to be in 2020 (\$50/m^2). Table 2.1 summarizes a number of future membrane and module manufacturing improvements and skid design optimizations that we see as a pathway to meeting cost targets. Each of these areas has the potential to make significant improvements in system costs. For example, during small pilot work, a number of membrane and module manufacturing bottlenecks were identified. Addressing these labor-intensive manufacturing steps, such as through use of a dedicated module production line with automated assembly steps, will result in a total estimated cost savings of \$70/m^2. None of these manufacturing improvements are particularly groundbreaking; they are items already implemented in the RO membrane industry, and could be applied to Polaris production with low risk.

With regard to skid design, we plan to move from the current 8-inch (20 cm)-diameter spiral modules (20 m^2) to 12-inch-diameter (30.5 cm) (50 m^2) modules for the multi-tube vessel that will be used in the vacuum step of the membrane process. This change produces an estimated \$30/m^2 savings in module housing cost. Because some commercial MTR membrane systems already use 12-inch modules (although not Polaris), this improvement is relatively "low-hanging fruit." We also plan to investigate the feasibility of using low-cost fiber-reinforced plastic or similar materials for membrane module housings. These materials are widely used in the RO membrane industry and should be applicable for use with low-pressure, non-flammable flue gas. The success of the low-cost plastic module components already validation tested on the 1 and 20 TPD systems gives us confidence that this approach has merit. Based on experience from the RO industry, we believe this change could result in savings of another $\$50/m^2$.

Table 2.1. Potential Membrane Skid Cost Reductions.

Cost Savings Category	Modification	Potential Savings ($\$/m^2$)
Membrane and module manufacturing	• Increase membrane roll size from 300 to 900 meters. • Automated module envelope assembly and glue application. • Dedicated module production line.	\$15 \$35 \$20
Skid design	• Impact of 12" (30.5 cm)-diameter modules on skid. • FRP manifolds, housings, and components.	\$30 \$50
	Total Potential Savings	~\$150

By following the pathway outlined above, we expect that Polaris membrane module skid costs can be reduced to near $\$50/m^2$. Further savings may be possible simply from economies of scale realized in going from the current 1 MW_e small pilot to future large pilots and demonstration systems. Based on this analysis, we feel very confident that membrane cost and performance targets are achievable.

3. FIELD TESTING OF A 1 TPD MEMBRANE SYSTEM

The main focus of this chapter is to discuss operational experience and performance results for the MTR 1 tonne/day (TPD) membrane CO_2 capture system that was tested at the National Carbon Capture Center (NCCC). The system was originally installed at NCCC in late 2011, and commissioned in early 2012. Overall, the system accumulated over 11,000 hours of operation until final retirement in August 2015. The objective of the field demonstration was to investigate membrane and module performance with coal-fired power plant flue gas, and to gain system operating experience that could be incorporated into the design of the larger 20 TPD small pilot system. In early 2014, this larger MTR small pilot system was delivered to NCCC, and collected experimental data in 2015. These results are reported in Chapter 4. Details of the 1 TPD system performance are also described in a recent publication.[11]

3.1 Operation of the 1 TPD System with a Dry Screw Compressor in 2012

As shown by the process diagram in Figure 3.1, the 1 TPD system removes CO_2 from flue gas in two steps. The first step uses cross-flow modules with a vacuum on the permeate for CO_2 enrichment, and the second step uses counter-current sweep modules to remove additional CO_2 in the feed gas to meet the overall capture target (~90%). Before entering membrane modules, the gas is compressed to 20 to 30 psig, to generate the necessary pressure ratio for parametric study of membranes installed on the system. For operation in 2012, an Atlas Copco dry screw compressor was used for feed gas compression.

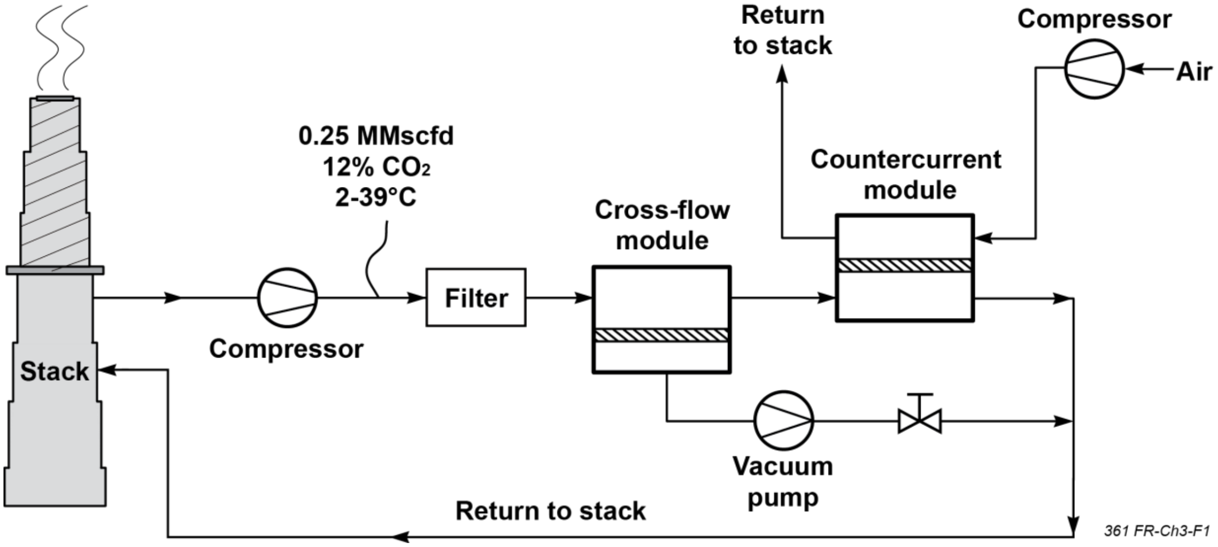


Figure 3.1. Process flow diagram of the 1 TPD membrane system that operated at NCCC from January 2012 to August 2015.

In 2012, flue gas outages (January to March) and compressor corrosion issues caused interruptions in the system operation. Most of the continuous system operation was obtained from April to June. Figure 3.2 shows the real time CO_2 content during operation from April 19 to June 26 for three gas streams of interest: the feed gas, the CO_2 enriched stream, and the CO_2 depleted stream. Air ingress was completely avoided after 300 hours of operation, and the CO_2 content in the feed gas stabilized at approximately 12%. From 300 to 1,000 hours of operation, the two-step membrane operation reduced the CO_2 content to 5% in the CO_2-depleted stream, indicating an overall capture rate of approximately 60%. The cross-flow modules enriched the CO_2 in the permeate by a factor of 6-8 times. With new sweep modules installed in the system in early June (after 1,000 hours of operation), the CO_2 content in the CO_2-depleted stream was further reduced to around 3%, indicating a CO_2 capture rate of 85%. Figure 3.3 shows the overall CO_2 removal rate of the membrane system over this same period.

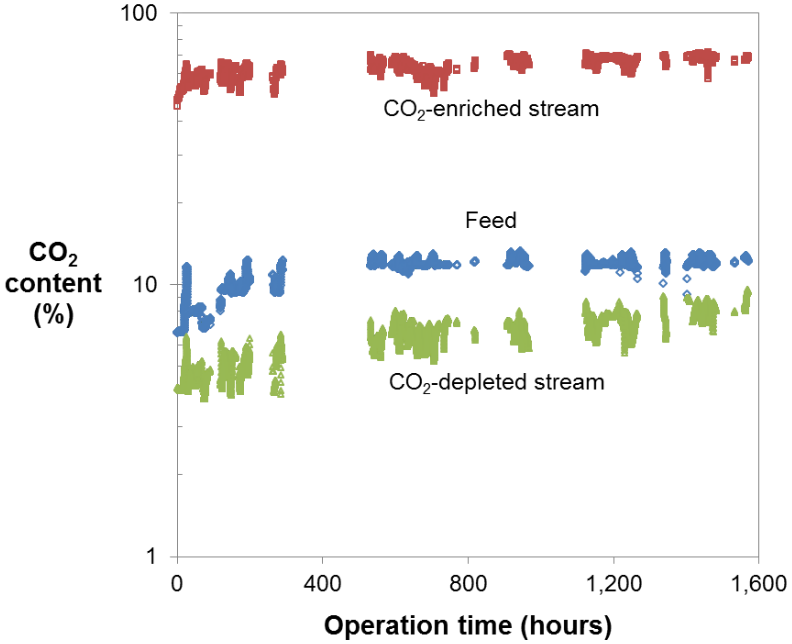


Figure 3.2. Real time CO_2 content of the feed gas, the CO_2-enriched stream and the CO_2-depleted stream during operation from April 19 to June 26, 2012.

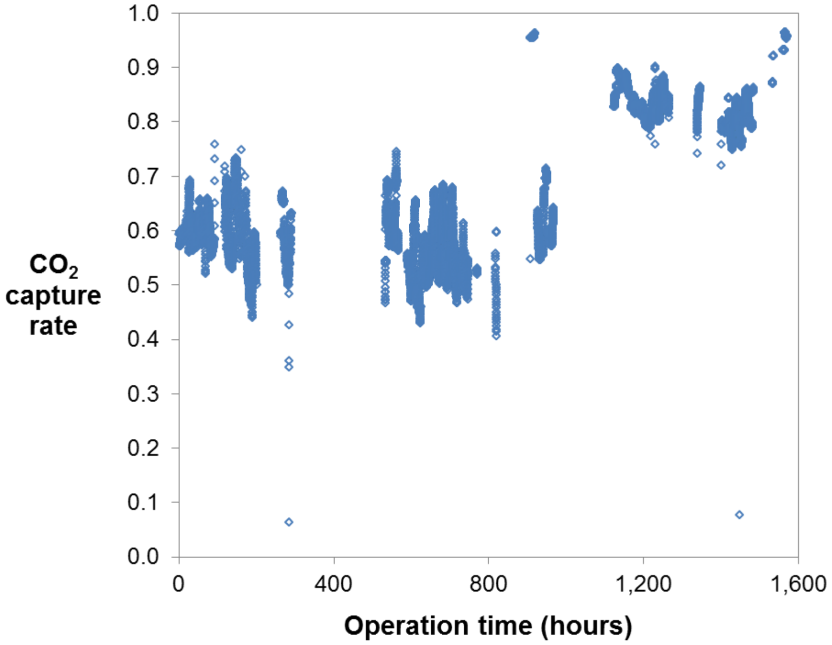


Figure 3.3. Overall CO_2 removal rate by the 1 TPD system during operation from April 19 to June 26, 2012.

Polaris modules were installed on the system prior to the system startup in January 2012. Modules 6114 and 6419 were returned to MTR in July for post-test analysis. As shown in Figure 3.4, no particles appear to have collected on either end of the module. Module 6419 was stripped and cut open. No deposition of particles was observed inside the membrane envelopes. Figure 3.5 shows pictures of membrane sheets and a feed spacer from module 6419 after it was cut open. Overall, the membranes were in excellent condition after about 1,500 hours of operation.

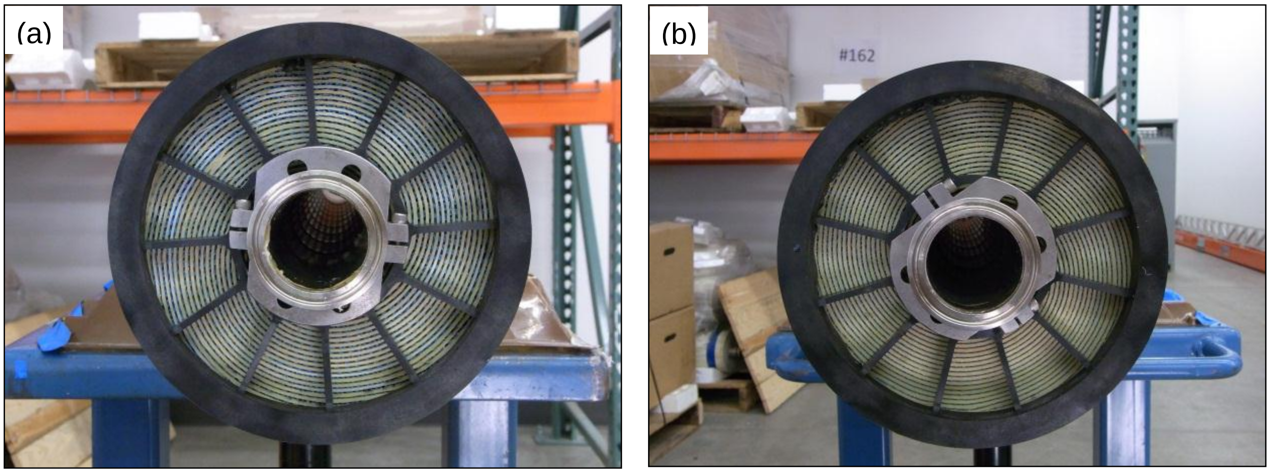


Figure 3.4. Pictures of feed gas inlet (a) and residue gas outlet (b) of module 6419. The module was tested on the 1 TPD system at NCCC from April to August of 2012.

Figure 3.5. Pictures of membrane sheets and feed spacer of the stripped module 6419. The module was tested on the 1 TPD system at NCCC from April to July of 2012.

While the membrane modules showed excellent durability during testing in 2012, the dry screw feed compressor did not perform nearly as well. A number of system shutdowns were traced to fouling and corrosion of the feed compressor. In particular, a power plant system upset in late July 2012 resulted in ammonium sulfate deposition on the compressor that caused system shutdown. After discussions with compressor equipment providers, it was decided to switch to a liquid ring compressor due to the greater tolerance of this compressor type to particulates and other contaminants in the flue gas feed.

3.2 Operation of the 1 TPD System using a Liquid Ring Compressor

In 2013, the Atlas Copco feed compressor was replaced with a Gardner Denver Nash liquid ring compressor, to improve the overall reliability of the system operation with flue gas. During the system modification, new cross-flow and sweep modules were rotated into the system. Figure 3.6 shows the cumulative module performance, with respect to CO_2 removal from regular coal-fired flue gas and CO_2 enrichment in the permeate streams for the modules that were tested on the system between December 14, 2012 and July 11, 2013. The cumulative run time of the modules was over 1,300 hours. The figure is divided into five sections, showing the module performance during each period of continuous operation. Overall, these modules again demonstrated stable performance at expected levels of separation, even after remaining idle in the membrane system several times during system repair/maintenance and flue gas/cooling water outages at NCCC. The fluctuation in module performance was mostly caused by the ambient temperature variation.

In July, modules that were tested on the 1 TPD system during this period were brought back to MTR for post-test analysis. Table 3.1 shows the operating history of these modules. Table 3.2 shows the pure-gas performance of these modules after testing on the 1 TPD system, relative to their original performance before the test. Both CO_2 permeance and CO_2/N_2 selectivity remained almost unchanged (within the error range of the pure gas module testing system), even after going through many cycles of system restart and shutdown, and staying idle in the membrane system for over three months during the flue gas outage at NCCC from February to April 2013.

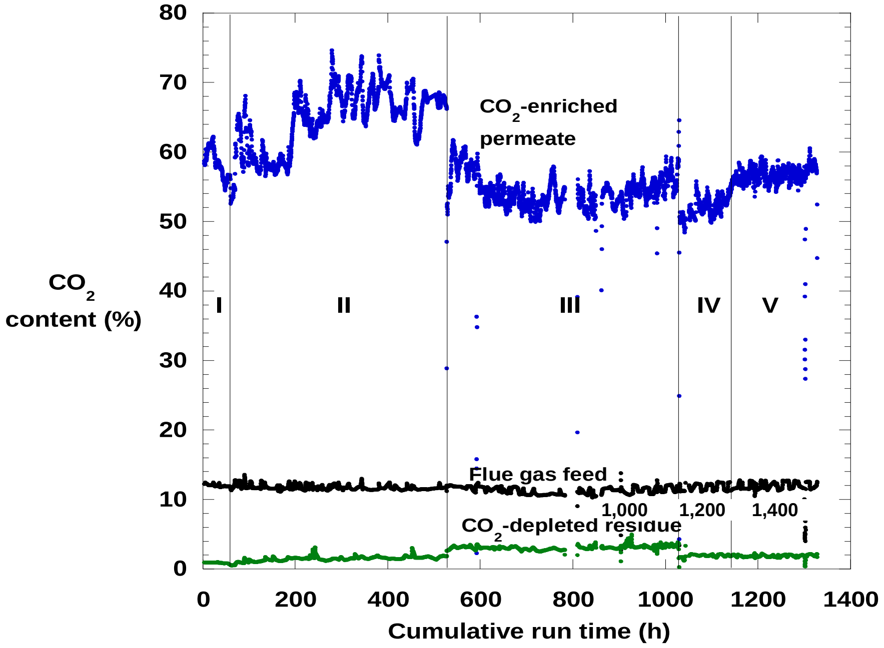


Figure 3.6. CO_2 content of gas streams entering and exiting the 1 TPD system as a function of time, when the system ran with undiluted coal fired flue gas feed. Period I: 12/14/2012 to 12/17/2012; Period II: 01/07/2013 to 01/28/2013; Period III: 05/03/2013 to 05/24/2013 (the system ran only with cross-flow modules during this period, due the leak issues with the sweep modules); Period IV: 06/20/2013 to 06/25/2013; Period V: 07/03/2013 to 07/11/2013.

Table 3.1. Operating History of the Cross-Flow Modules that were Tested on the 1 TPD System from December 14, 2012 to July 17, 2013.

Time	Operating Status
December 13, 2012	Module installation
December 14 - 17, 2012	In operation with flue gas
December 18, 2012 - January 8, 2013	Standing idle in the 1 TPD system
January 9 - January 28, 2013	In operation with flue gas
January 29 - May 2, 2013	Standing idle in the 1 TPD system
May 3 - May 24, 2013	In operation with flue gas
May 25 - June 19, 2013	Standing idle in the 1 TPD system
June 20 - 25, 2013	In operation with flue gas
June 26 - July 2, 2013	Standing idle in the 1 TPD system
July 3 - July 17, 2013	In operation with flue gas

Table 3.2. Relative Performance of Cross-Flow Modules Before and After Testing on the 1 TPD System from December 2012 to July 2013.

Module	Normalized CO_2 Permeance	Normalized CO_2/N_2 Selectivity
6704	87%	94%
6706	111%	130%

On July 11, 2013, NCCC started diluting flue gas with air, to simulate the CO_2 content in the flue gas for a natural gas combined cycle (NGCC) power plant. At the same time, new modules were installed in the 1 TPD system for performance validation. The new modules were made of membranes that were produced using an optimized membrane fabrication process. The process combines multiple processing steps into one single step, which allowed for reduction in processing time, as well as labor and materials costs. This translated into an overall saving of 15% in module production.

Figure 3.7 shows the cumulative module performance, with respect to CO_2 removal from air-diluted coal-fired flue gas and CO_2 enrichment in the permeate streams for the modules that were tested with the air-diluted flue gas from July to October 2013. The system removed approximately 80% of the CO_2 from flue gas, and enriched it by a factor of 7-8 in the permeate stream. The 1 TPD system was originally designed to capture ~90% CO_2 from regular coal-fired flue gas at a total flow rate of approximately 550 lb/hr. Diluting the flue gas with air caused the CO_2 content to drop from 12% to 4%, resulting in a significant reduction in the CO_2 partial pressure of the feed gas. Separation of CO_2 by membrane technology is primarily driven by the differential in the CO_2 partial pressure. Therefore, without changing the feed gas flow rate and other operating conditions, less CO_2 (percentage-basis and mass-basis) is removed from the feed gas when air-diluted flue gas is used. Nevertheless, the system still achieved more than 80% CO_2 capture and an almost 8-fold CO_2 enrichment in the permeate stream.

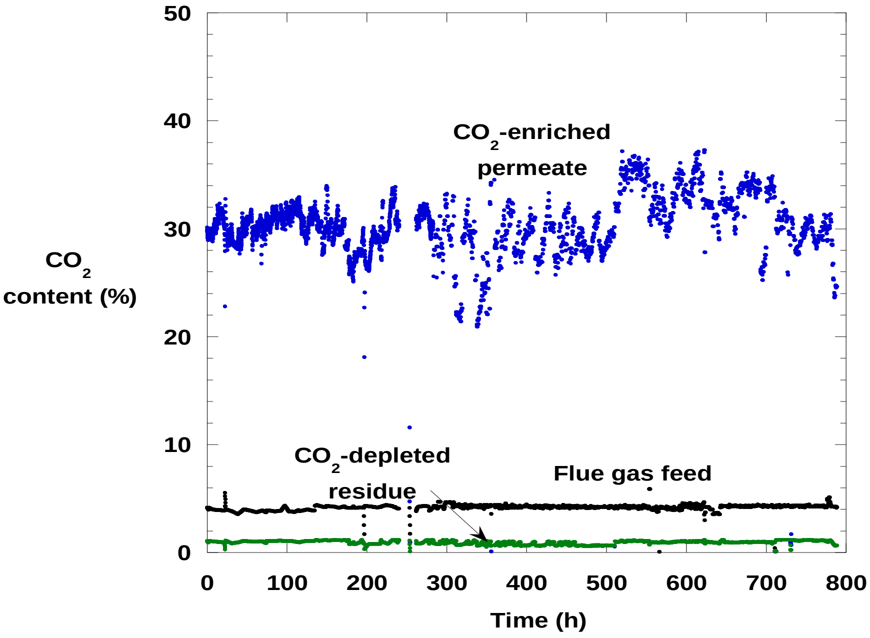


Figure 3.7. CO_2 content of gas streams entering and exiting the 1 TPD system as a function of time in latter half of 2013; the system ran primarily with air-diluted coal-fired flue gas.

If we were able to test the full MTR capture process, CO_2 recycle by sweep would increase the feed CO_2 content from 4% to 15-20%. Based on the module performance shown in Figure 3.7, simulations indicate that the full-process CO_2 enriched permeate would contain >80% CO_2, and be ready for final purification and compression.

3.3 Testing the Second Generation (Gen-2) Polaris Membrane on the 1 TPD System

In light of the smooth operation of the 1 TPD system in late 2013, testing of the system in 2014 focused on validating the performance improvement of Polaris membranes in the field. The 1 TPD system has two pressure vessels in parallel in the cross-flow step, which allows the testing of multiple cross-flow modules. In January 2014, we rotated modules made of advanced Polaris membranes into one of the vessels for the validation of membrane performance improvement. Meanwhile, old modules that were tested in 2013 were kept in the other vessel for long-term stability monitoring. Table 3.3 lists the module identification information for the cross-flow modules that were tested on the 1 TPD system in this period.

Table 3.3. Cross-Flow Modules Being Tested on the 1 TPD System in early 2014.

Module #	Membrane	Testing Period
6706	Base-case Polaris ($0.5/m^2 gpu)*	I: 12/2012 to 07/2013 II: Since 01/2014
7143	Low-cost Polaris ($0.2/m^2 gpu)	I: 07/2013 to 10/2013 II: Since 01/2014
7297	Advanced Polaris ($0.13/m^2 gpu)	I: Since 01/2014
7298	Advanced Polaris ($0.13/m^2 gpu)	I: Since 01/2014

*Permeance normalized membrane cost

Figure 3.8 shows the system performance during operation with coal flue gas in 2014. When the system ran with full capacity (during the first 100 hours and after 300 hours shown in Figure 3.8, it captured over 80% of the CO_2 from flue gas, and enriched the CO_2 in the permeate stream by a factor of 5.

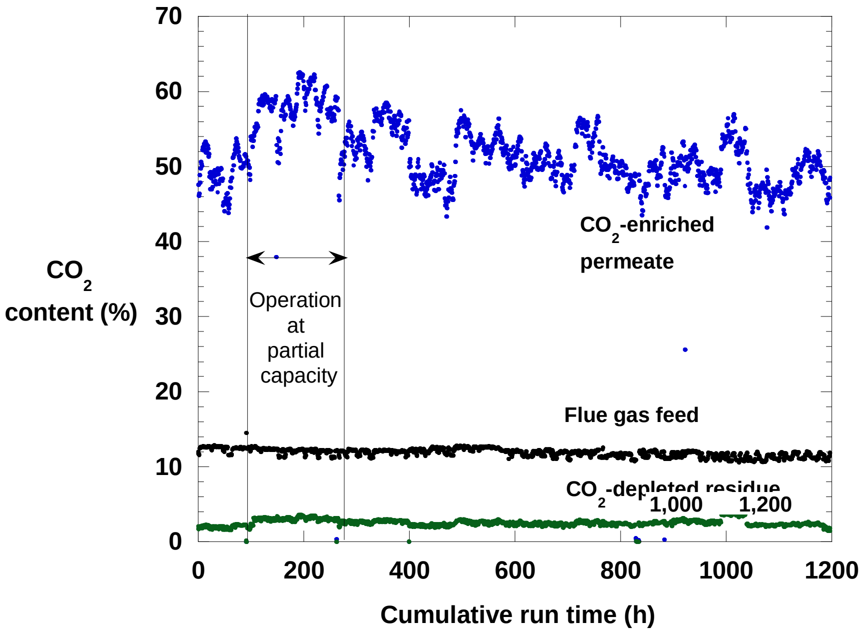


Figure 3.8. CO_2 content of gas streams entering and exiting the 1 TPD system as a function of time during operation in early 2014.

The performance data (conditions and composition of each key stream) of the 1 TPD system were recorded for the main streams entering and exiting each separation step, and each step has two pressure vessels in parallel. As a result, when the system ran at full capacity, only the average module performance could be obtained. To measure the module performance in each individual pressure vessel, we ran the system with only one vessel open in the cross-flow step after the first 100 hours of operation in 2014, alternating module use as indicated in the figure. As shown in Figure 3.8, when the capacity was reduced to 50% in the cross-flow step, the system was still able to capture over 70% of the CO_2. Under this condition, the sweep step worked at higher removal efficiency to make up part of the capacity loss in the cross-flow step, due to the higher CO_2 content in the gas fed to the sweep step.

Modules 7297 and 7298 were made using Gen-2 Polaris membranes that have at least 50% higher CO_2 permeance than that of the base-case Polaris membranes. Therefore, the modules were expected to have higher CO_2 removal capacity than the base-case modules that were tested in 2013. This is confirmed by the results shown in Figure 3.9. Under similar operating conditions, the Gen-2 module showed 60% higher CO_2 removal rate than the base-case module.

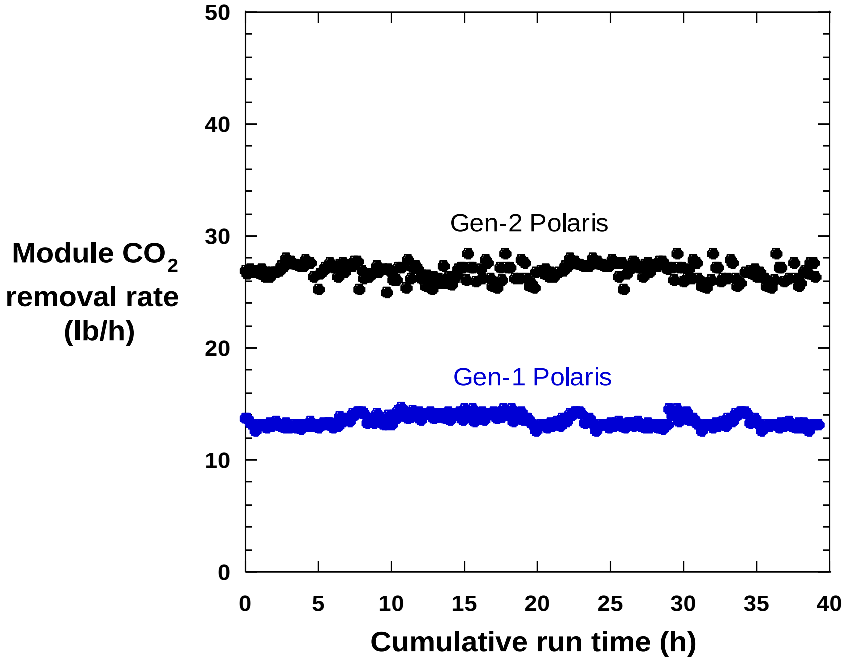


Figure 3.9. Comparison of the individual module CO_2 removal rate of a Gen-2 Polaris module and a base-case Polaris module. The results were calculated based on the system performance data recorded during February 11-13, 2014.

3.4 Operation of the 1 TPD System at NCCC in 2015

The 1 TPD system started up in January 2015 with high-flux Polaris cross-flow modules in the first step, and sweep modules with high selectivity Polaris membrane in the second step. This combination of modules had not previously been tested. The objectives were to determine module performance with coal generated flue gas in the field, and to test if these modules provide stable performance over time.

The system was loaded with high flux Polaris cross-flow modules manufactured in November 2014. Half the modules went into each cross-flow housing to maximize membrane performance by supplying a sufficient quantity of flue gas. Previous high flux modules had been starved for gas limiting the observed gains expected in system performance due to the high flux. Sweep modules with higher CO_2/N_2 selectivity retained from the previous campaign were in the second-step sweep housing. Higher selectivity reduces the changes in the sweep air composition which exits the modules. The test system was started up on January 14, 2015.

Figure 3.10 shows the initial weeks of system performance during operation with flue gas feed. Flue gas containing 12% CO_2 was enriched to better than 60% CO_2 in the first-step permeate, with the outlet residue gas from the second-step sweep modules at 2.8% to 4.0% CO_2. The ambient temperatures during January trended downward as the run progressed which promotes higher membrane selectivity and steadily improved the %CO_2 over time in the first-step permeate stream. The system was run with air on January 18-20 and 27-31 when flue gas was unavailable. It is important that both the cross-flow and sweep modules withstand these changes in process conditions as the power plant goes on and off-line.

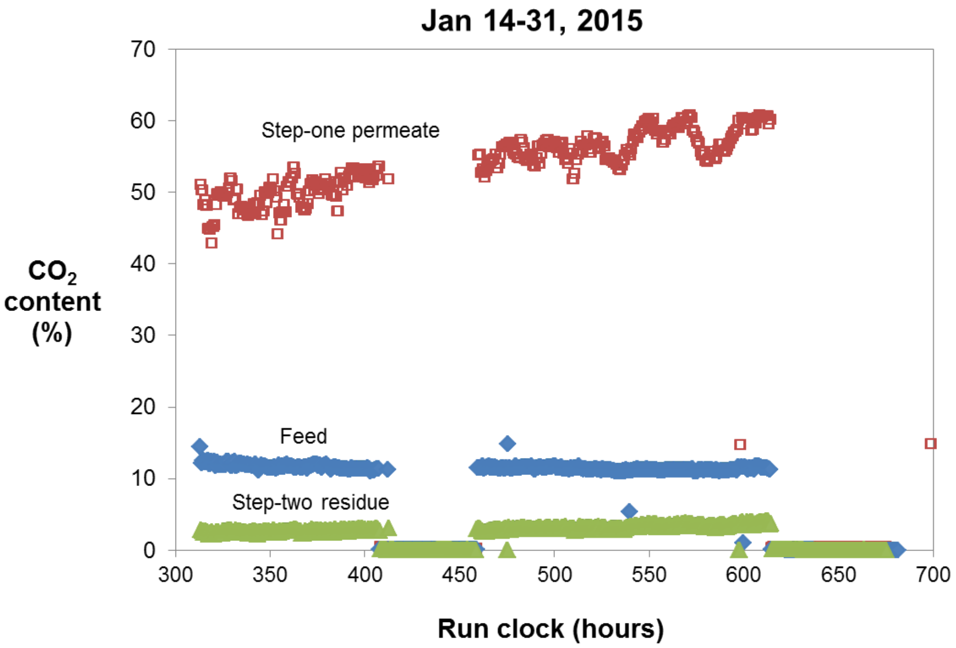


Figure 3.10. CO_2 content of gas streams entering and exiting the 1 TPD system as a function of time during system operation in January 2015.

One of the remarkable features observed in running full-scale modules was that even though the system was running under ambient conditions outdoors at NCCC, some of the observed outputs (such as residual CO_2 content) were consistent from day-to-day in both the first and second steps. This behavior reflects that the membrane system, when in operation, demonstrates self-leveling features. "Self-leveling" results from the concept that in a complex system with multiple linked variables, some of the performance functions respond in opposite directions, leading to more constant performance than might first be expected.

The performance of the test system from January to July 2015 is shown in Figure 3.11. For the operating periods with flue gas the feed is about 12% CO_2. The step-one permeate ranged from 50 to 70% CO_2 representing an enrichment factor of 4 to 6. The system generates a treated flue gas containing between 2 to 5% CO_2 as the second-step residue stream. The restart of operations in June 2015 after an extended shutdown showed that stable operations continued.

Previous operations during NCCC campaign PO-2 ceased on March 26, so with the restart on June 11, the system downtime was over 1,700 hours. Figures 3.12 and 3.13 detail just the operations in June and July. Upon restart, the feed rate to the system approached 200 kscfd, which increased the second-step residue to just over 5% CO_2. Lowering the feed rate to 150 kscfd returned the second-step residue to under 4% CO_2. In mid-June the feed temperature to the membrane system went up to about 100 °F. This increases the permeation rate, which increases stage cut, with the net result of lower CO_2 content in the first-step permeate and lower CO_2 in the second-step residue.

The period of operations in February was defined by the external winter weather with feed streams cooling from around 80 °F at the start to 55 °F. Lower feed temperature means lower permeate flux in both stages, lower stage cut in step one, higher %CO_2 in the first-step permeate, higher %CO_2 in the first step residue, and higher %CO_2 in the second step residue (the treated flue gas). The fairly steady trend to cooler temperature meant a steady shift in observed CO_2 content as the modules responded over time. Previous extended trials with this system have shown similar impacts due to operating temperature and these 2015 trials showed results consistent with the earlier trials.

For a brief period from 833 to 855 hours, the feed switched from pulverized coal flue gas (PCFG) to simulated flue gas from natural gas at 4% CO_2 content. The permeate from step one falls to about 40% CO_2 and this represents an enrichment factor of about 10. With the lower CO_2 content in the feed compared to coal flue gas, the second step residue is now reduced to less than 2% CO_2. Upon returning to 12% CO_2 feed, the performance recovered with CO_2 in the first-step permeate returning to about 60% CO_2.

At 4,020 run hours the feed rate was lowered to about 100 kscfd, which increases stage-cut and generates immediate changes in first-step permeate and second-step residue, both again to lower CO_2 content. At the end of June the sweep rate in the second step was both lowered and raised by 15 kscfd. This changes both the pressure drop across the sweep-side of the module, and the dilution rate of the permeate CO_2 for the second step. Increasing the sweep rate also had a small but measurable improvement in the amount of CO_2 captured by the sweep modules. But overall, the sweep flow rate has much less impact on system performance, compared to the feed flow rate or temperature.

Details of operations at the end of the run are shown in Figures 3.12 and 3.13. Again, the first step is taking a feed stream of 12% CO_2 to over 60% CO_2 in the permeate. The second step residue ranges from 2-7% CO_2. The key parameters governing system performance are the feed flow rate, the feed gas temperature, and the sweep flow rate. Intentional changes were made to vary the sweep flow rates between 20 to 65 kscfd, which impacts the step two residue CO_2 content, and the composition of the sweep-out gas.

Data collection ceased on July 17 at 4,622 hours signaling the end of system operation at NCCC. The system achieved over six months of operations with this combination of cross-flow and sweep modules. The modules continued to show stable operations even with the excursions in feed temperature and flow rates.

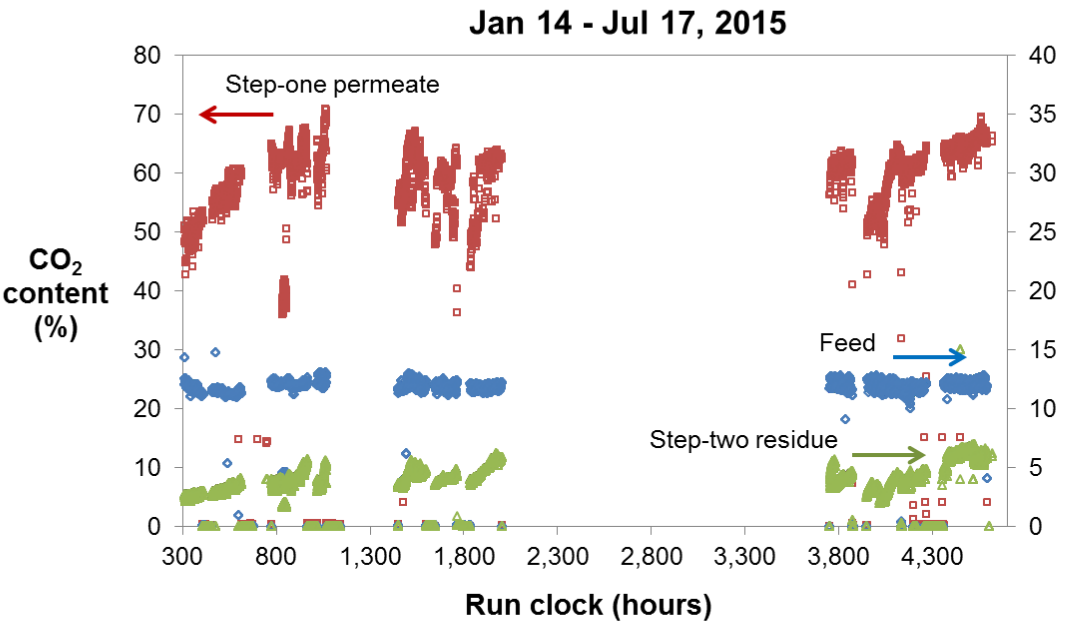


Figure 3.11. CO_2 content of both the feed and permeate gas streams for the first step of the 1 TPD system and the residue of the second step from operations over January to July 2015.

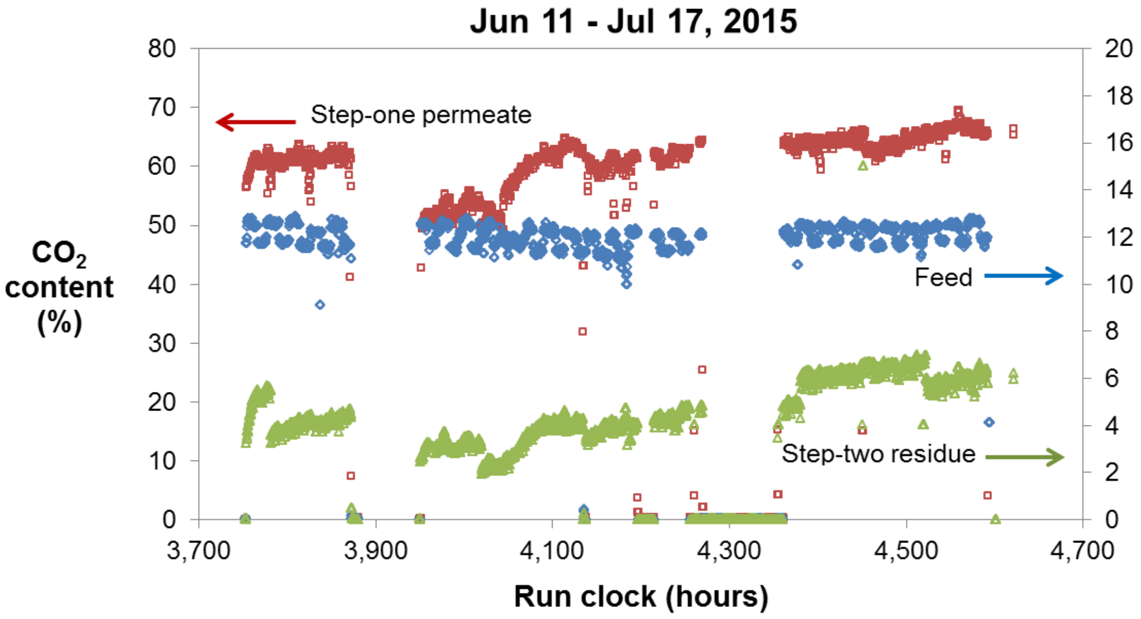


Figure 3.12. CO_2 content of both the feed and permeate gas streams for the first step of the 1 TPD system and the residue of the second step from operations for June to July 2015.

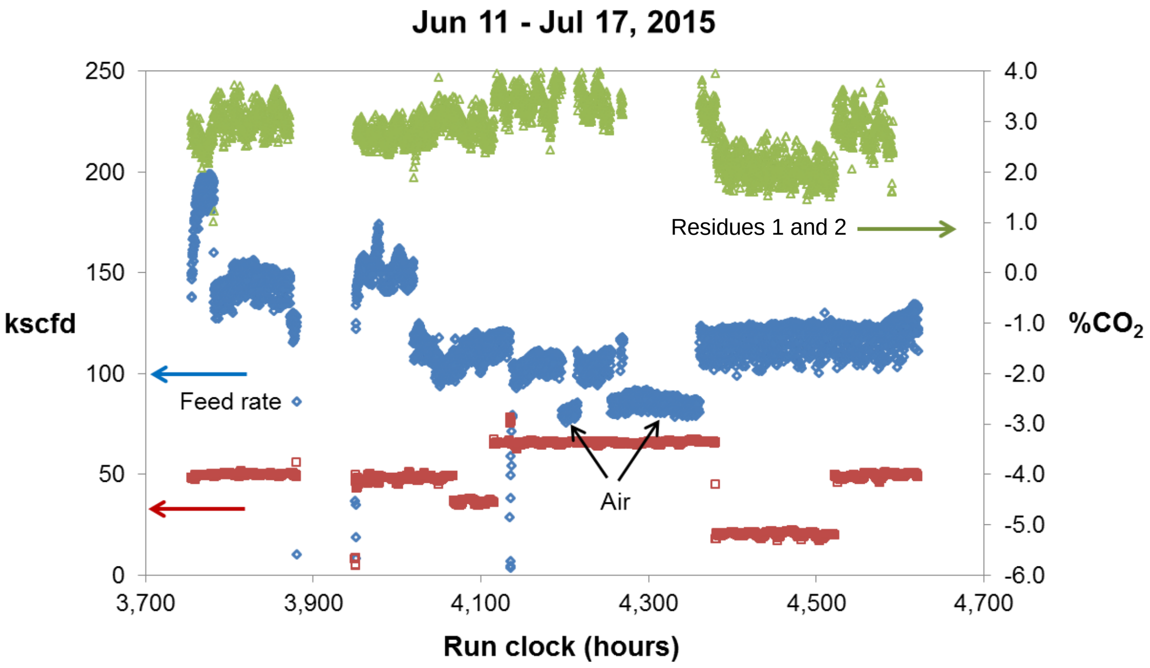


Figure 3.13. Key control parameters for 1 TPD system of feed gas temperature, feed rate to step one, and sweep in rate for step two at end of run.

3.5 Lessons Learned from System Operation

3.5.1. Dry Screw Compressor

The selection of the dry-screw compressor as part of the 1 TPD system was based on the previous operation experience at APS Cholla power plant. The system tested at Cholla in 2010 used an oil-flooded screw compressor that suffered from severe corrosion issues caused by water condensation from flue gas. This issue can be avoided with a dry-screw compressor, because it does not use any sealing fluid to generate high pressure ratio (inlet/outlet), and usually operates at a very high temperature (360-380 °F). In addition, the casing and screw of the Atlas Copco compressor were coated with Teflon® and epoxy, respectively, in an attempt to protect the compressor from corrosion due to atmospheric humidity. Early operation experience in 2012 showed that the dry screw compressor had better resistance to corrosion than the oil flooded compressor.

However, after three-months of relatively smooth operation, corrosion issues started to surface. Rust and black particles were found inside the dry screw compressor. Figure 3.14 shows pictures of some components after they were dissembled from the compressor in August 2012. It appeared that the corrosion–resistant coating layer did not protect against long-term exposure to flue gas. Presumably, after the coating layer was worn out, the compressor elements were in direct contact with flue gas, and corrosion issues started to occur gradually. In addition, without the sealing fluid, the compressor has a tight-fit screw-casing design, which reduces the tolerance to particles, dust and rust. These findings indicate that a dry screw compressor is not suitable for operation with flue gas.

Figure 3.14. Pictures of the feed compressor components showing particulate buildup and corrosion.

3.5.2. Liquid Ring Compressor

In light of re-occurring feed compressor issues, we decided to replace the Atlas Copco dry screw compressor with a Gardner Denver Nash liquid ring compressor. Liquid ring compressors are used in the petrochemical industry to handle toxic, corrosive, and explosive gases, and can be ordered with stainless steel internals. With a liquid ring as the sealing media, these compressors have a greater tolerance for solids in the feed. Nash provided a one-year warranty, a list of references, and installation/operating support at NCCC.

Figure 3.15 shows the cumulative operating time of the 1 TPD system in 2012 and 2013 when flue gas was available at NCCC. When operated with the liquid ring compressor in 2013, the overall reliability of the 1 TPD system was significantly improved. Some interruptions in the early operation were caused by mechanical issues due to the inappropriate system assembling and handling, but not related to working with flue gas. The compressor was thoroughly inspected by the Nash representative after one month of operation with flue gas. Nash confirmed that all the key components of the compressor were in "good condition." No pitting or erosion was observed. Figure 3.16 shows pictures of key components when the system was dissembled during the inspection.

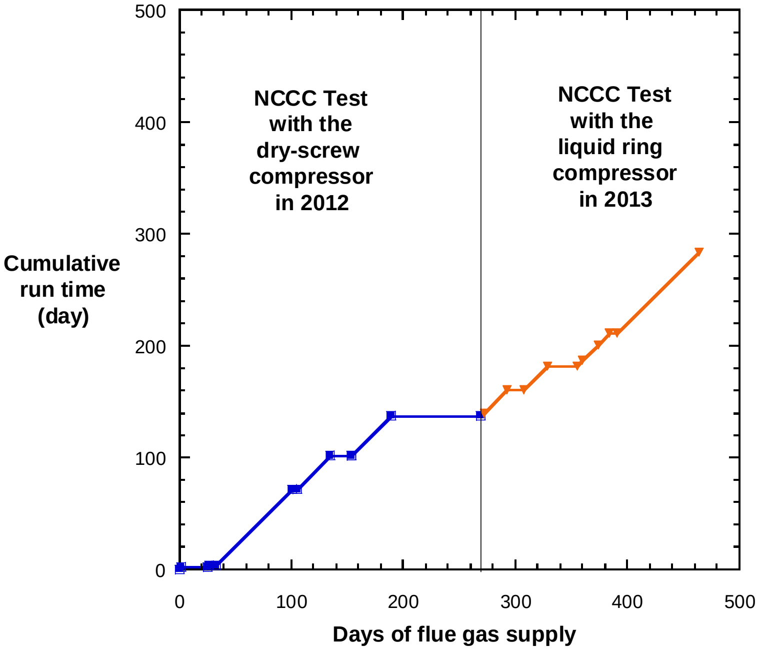


Figure 3.15. Cumulative run time of the 1 TPD system at NCCC when flue gas was available.

Figure 3.16. Pictures of key compressor components after the liquid ring compressor operated with flue gas for over one month at NCCC in early 2013.

3.5.3. System Purging During Shutdowns

The system performance data and post-test module analysis discussed in previous sections indicate that, when the system was in operation, membrane modules had stable performance against real power plant flue gas. In 2012, after the Atlas Copco compressor was taken offline due to corrosion issues, the system ran with the NCCC blower upstream of the flue gas delivery line, to collect module lifetime data by simply exposing membrane modules to flue gas continuously. Figure 3.17 shows the real-time feed gas pressure generated by the blower, the CO_2 content (wt%) in the feed gas, and the CO_2 content (wt%) in the residue gas leaving the cross-flow modules on the 1 TPD system. Due to the lack of driving force, only about 30% of the flue gas CO_2 is removed by the cross-flow modules. Nevertheless, the relatively steady CO_2 content in the residue gas indicates stable performance of the membrane modules.

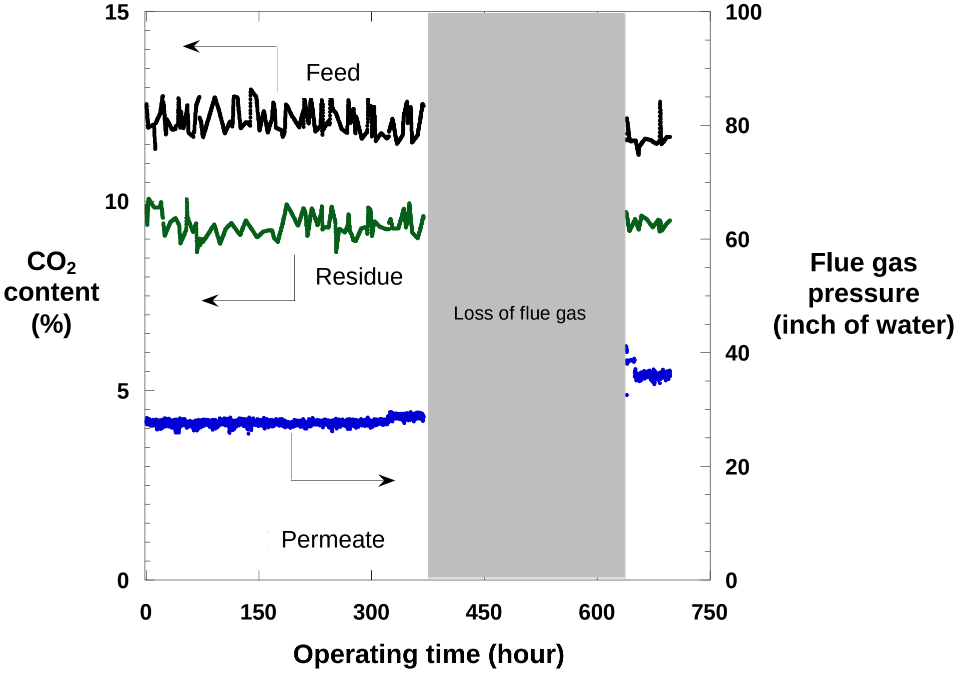


Figure 3.17. 1 TPD system performance (August and September 2012) when operating with a feed blower and permeate vacuum pump.

The exposure test ended in September 2012, and the modules were stored inside the pressure vessel until December 2012. They were then returned to MTR. When tested at MTR in December, they showed low flux and selectivity.

Module 6115 was cut open for a more detailed examination. All of the membrane leaves within the module appeared to be filled with liquid. Liquid was also observed as micro-droplets dispersed on the membrane surface. Figure 3.18 shows a picture of a membrane leaf from Module 6115. This liquid was collected with a syringe. Indicator paper shows the sample is strongly acidic in the range of 0-2 pH. A drop of this liquid was placed on the stage of a Bruker Tensor 27 Infrared spectrometer, and the reflectance spectrum generated is shown in Figure 3.19. The spectrum of fresh water was also recorded and shown in Figure 3.19 for comparison. There is an excellent match between the spectra, indicating that the majority of the unknown liquid within Module 6115 is water.

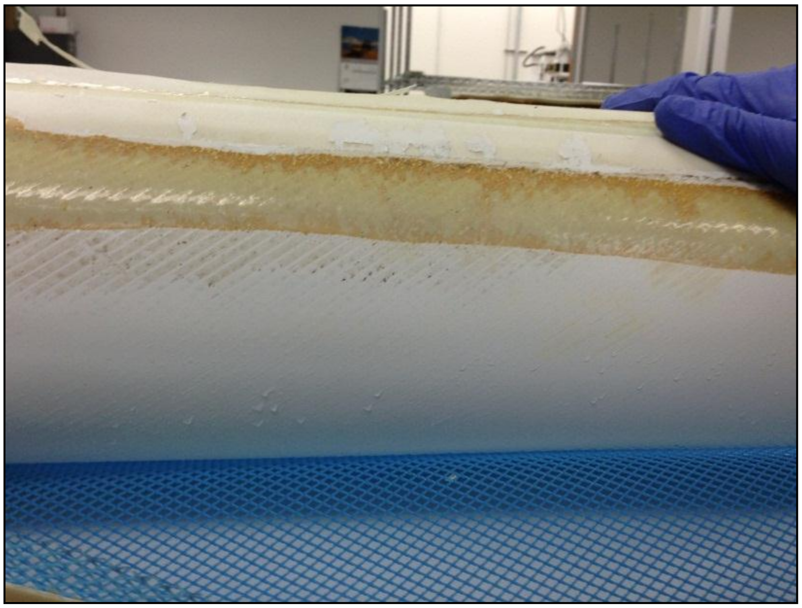


Figure 3.18. Photo of a membrane leaf from Module 6115. The module was tested intermittently on the 1 TPD system during the period from November 2011 to September 2012. It was removed from the system in December 2012.

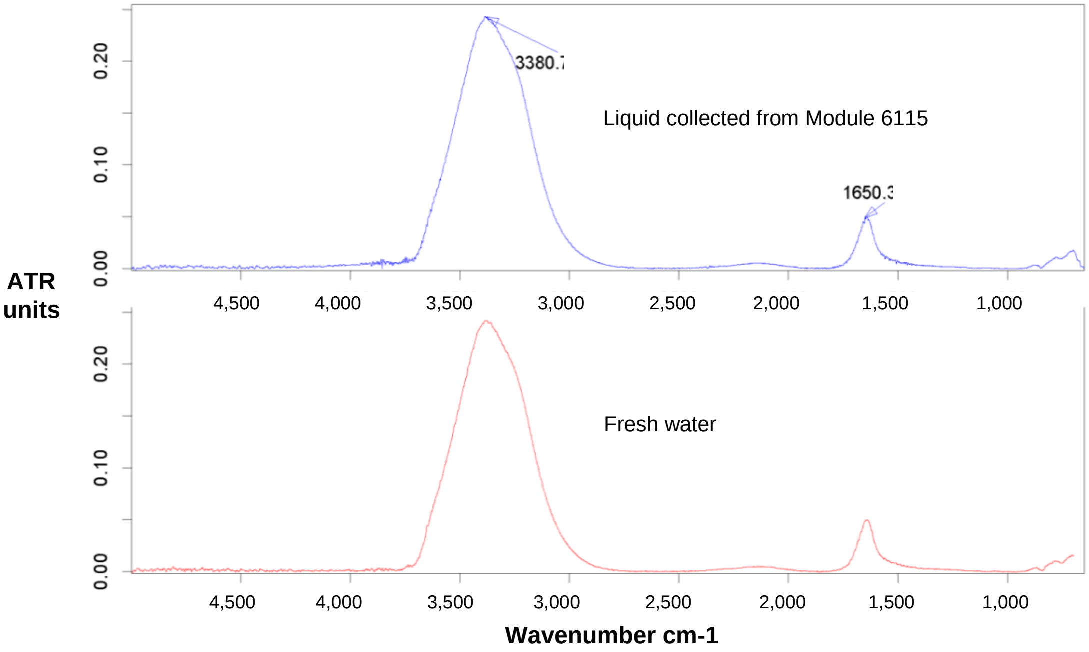


Figure 3.19. Reflectance infrared spectra of unknown liquid from Module 6115 (upper) and water (lower).

Evaporation of the liquid in the module left a visible white residue. Previous analyses of particles deposited on the outlet of the Atlas Copco compressor indicate that the flue gas at NCCC contains ammonium sulfate/bisulfate, which are byproducts of the selective catalytic reduction (SCR) unit for NO_x removal. Given this information, the IR spectra of the white residue and ammonium bisulfate (obtained from Sigma-Aldrich) were compared as shown in Figure 3.20. There is a good match between the features of the two spectra, suggesting that a majority of the solids present in Module 6115 are a mixture of ammonium sulfate/bisulfate. Because the white residue particles appear to be present in significant quantities, either on or within the active membrane layer, they could hurt permeation properties if they collect over time and are present in large quantities in the membrane system. With this lesson learned, the system was purged with air after each shutdown in later operations, and water condensation within the modules was avoided. The stable module performance shown in Table 3.2 indicates purging the membrane system with air can effectively prevent membrane fouling caused by water condensation and salt deposition.

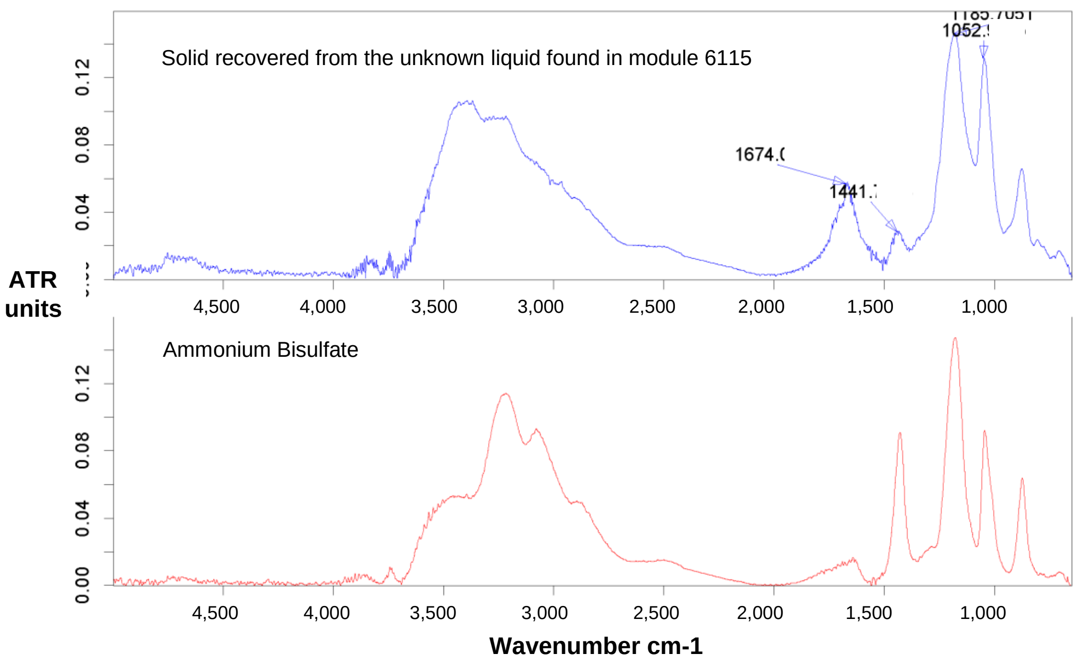


Figure 3.20. Reflectance infrared spectra of solids recovered by evaporation of water from Module 6115 (upper) compared with ammonium bisulfate evaporated from water (lower).

3.6 Effect of Different Module Configurations on Performance

One of the parameters that could be changed during this field test was the membrane modules themselves. Over this test period, Polaris membranes with different permeation characteristics, as well as modules with different packing configurations were studied. Table 3.4 reports data for six cases in which different modules were tested during the 2013 and 2014 campaigns. Cases 1 and 2 were taken from October 2013 trials, Cases 3 and 4 correspond to data taken before and after the test system was switched to operation with half the cross-flow modules in early 2014, and Cases 5 and 6 are data points from two time periods when only half the cross-flow modules were in operation in 2014.

Table 3.4. Average Performance Values for the Membrane Process First Step over Six Selected Time Periods with Steady State Operations at NCCC.

Case	Modules Used	Flow Rate (L/Min)		Temperature (°F)	Vacuum Pressure (psia)	Stage-Cut (%)	CO_2 Concentration (%)		
		Feed	Permeate				Feed	Permeate	Residue
1	A	2104	360	68.4	1.5	17.1	12.09	58.10	4.64
2	A	2851	397	70.3	1.7	13.9	11.95	61.05	5.65
3	B + C	2902	558	68.5	2.4	19.2	12.47	49.25	4.27
4	B + C	3066	533	67.7	2.2	17.4	12.35	50.84	4.65
5	only B	2768	352	62.9	1.6	12.7	12.12	59.20	6.37
6	only C	2760	421	73.4	1.9	15.3	12.00	54.26	5.55

Cases 1 and 2 from October 2013 used baseline Polaris modules. Cases 3 to 6 from 2014 used the higher permeance Gen-2 Polaris modules. As a result, Case 6, which uses only half the number of modules, shows a higher permeate rate than Cases 1 and 2 operating at full module capacity. In relative terms, the permeance of the modules follow the order: C > B >> A. This is the same conclusion derived from Figure 3.9 in that the second generation Polaris membranes capture higher amounts of CO_2.

Figure 3.21 uses the six cases to show that with increasing stage-cut, both the CO_2 content in the first-step permeate and residue decrease. At low stage-cuts, the average CO_2 concentration on the feed side is higher, so the average permeate concentration is also higher. As the stage-cut increases, the CO_2 in the residue decreases (and CO_2 capture rate increases), but CO_2 purity in the permeate decreases because there is no longer sufficient CO_2 in the feed to sustain a high permeate CO_2 concentration. This is an example of the typical recovery/purity tradeoff observed for membrane systems. You can have either high CO_2 purity or high recovery (capture rate) in the permeate, but not both without multiple stages or steps.

The Figure 3.21 data also show that the lower flux modules used in Cases 1 and 2 have higher CO_2/N_2 selectivity than the modules used for Cases 3-6. This is illustrated by the observation that, at the same stage-cut, the CO_2 permeate concentrations for Cases 1 and 2 are higher in comparison with the trend line for Cases 3-6. While the module selectivity has some impact on CO_2 purity, it has a lesser effect on residue CO_2 concentration (a proxy for CO_2 removal). As stage-cut increases, the residue CO_2 concentrations decrease, with all six cases positioned near a single trend line.

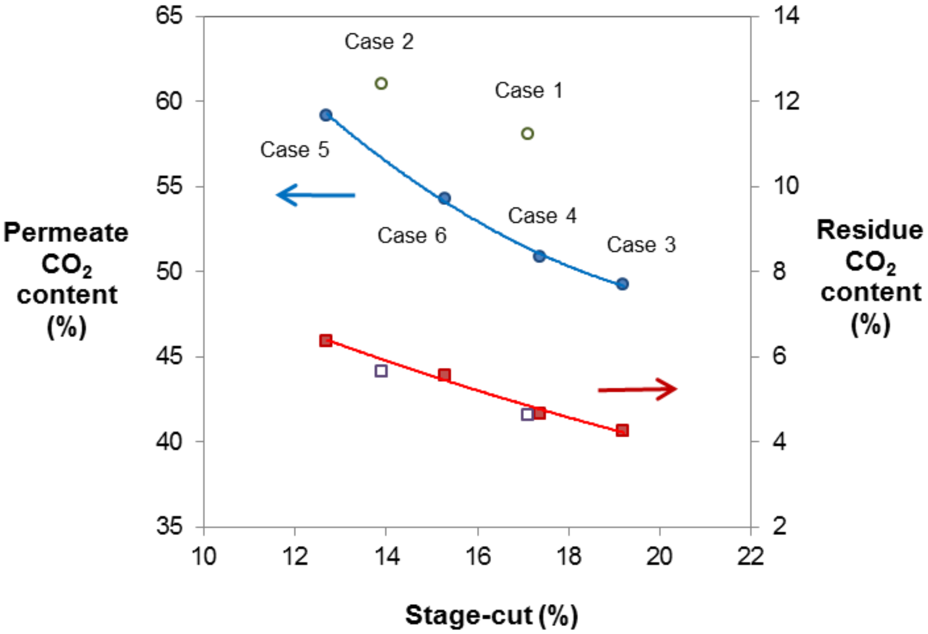


Figure 3.21. Average stream content data for six selected cases with 12% CO_2 flue gas illustrate that increasing stage-cut decreases the CO_2 concentration of both the first-step permeate and residue (see Table 3.4 for case details).

During the course of these module tests, an example of the impact of system design on performance was observed. Figure 3.22 shows the vacuum generated in the first step permeate as a function of permeate flow rate for the six cases described in Table 3.1. As permeate flow rates increase due to installation of higher flux modules, the first-step permeate pressure increases (i.e., a weaker vacuum is generated). Weaker vacuum means higher partial pressures of CO_2 in the permeate channel and thus lower pressure ratio, which limits the potential enrichment factor. These two factors of flux and vacuum work against each other and provide self-leveling performance. Because of this constraint, placing high-flux membranes in this test system did not greatly improve overall performance, even though the modules installed in 2014 provided better CO_2 permeance. This is an example of how engineering design and balance-of-plant equipment can impact membrane system performance. To take full advantage of the higher permeance modules, a higher capacity vacuum pump would need to be installed, and additional flue gas would need to be supplied along with a larger compressor in order to maintain the ~90% capture rate.

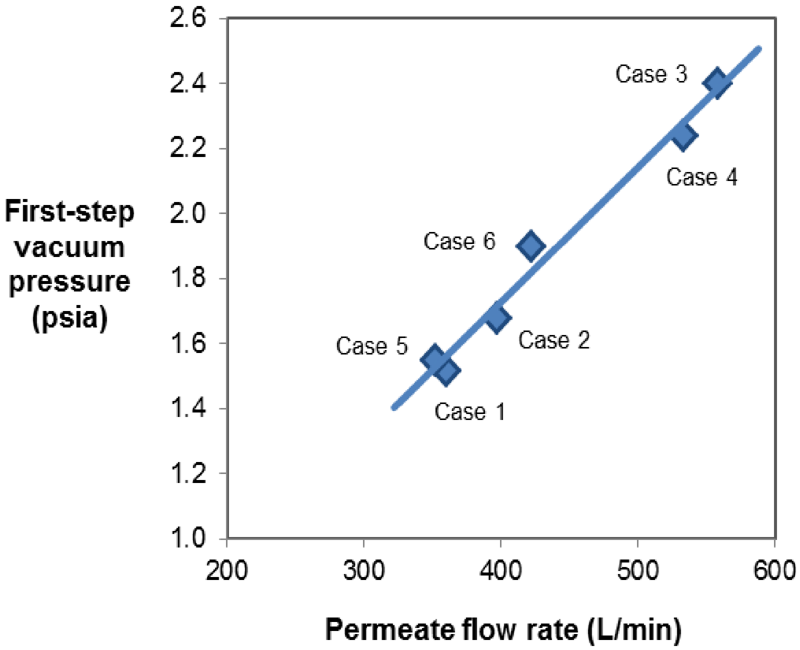


Figure 3.22. Average performance data from six selected cases illustrate that higher first-step permeate flow rates weaken (i.e., increase) the available first-step vacuum pressure (see Table 3.4 for case details).

3.7 Effect of System Temperature on Performance

The average system temperatures reported in Table 3.4 are in the range of 63-73 °F. Figure 3.23 uses data from January 2013 to show that the temperature swings over each day can be much larger. These data are from Polaris modules that operated between December 2012 and July 2013, providing over 1,200 hours of run time, but including some long periods of shutdown – up to three months. The largest direct impact of temperature is on the first-step permeate flow rate, which correlates strongly as shown in Figure 3.24(a). Here, the data are plotted as ℓn (permeate rate) versus inverse temperature, fashioned to illustrate the Arrhenius temperature dependence of permeation in the membrane modules. Using the nondispersive infrared (NDIR) sensor data for CO_2 and O_2 concentrations, the individual temperature responses for the three main flue gas constituents (CO_2, O_2, and N_2) can be determined as shown in Figure 3.24(b). The lowest curve is for O_2 since it is present at the lowest feed concentration (7%) although the inherent permeability is more than 2 times that of N_2. From the slope of these curves, the order of the mixed-gas permeation activation energies is as follows: $CO_2 << O_2 < N_2$. This order is consistent with Polaris being a solubility selective polymer membrane. While these temperature effects are to be expected for polymer membranes, it is rare that they are reported for full-scale modules operating in field trials.

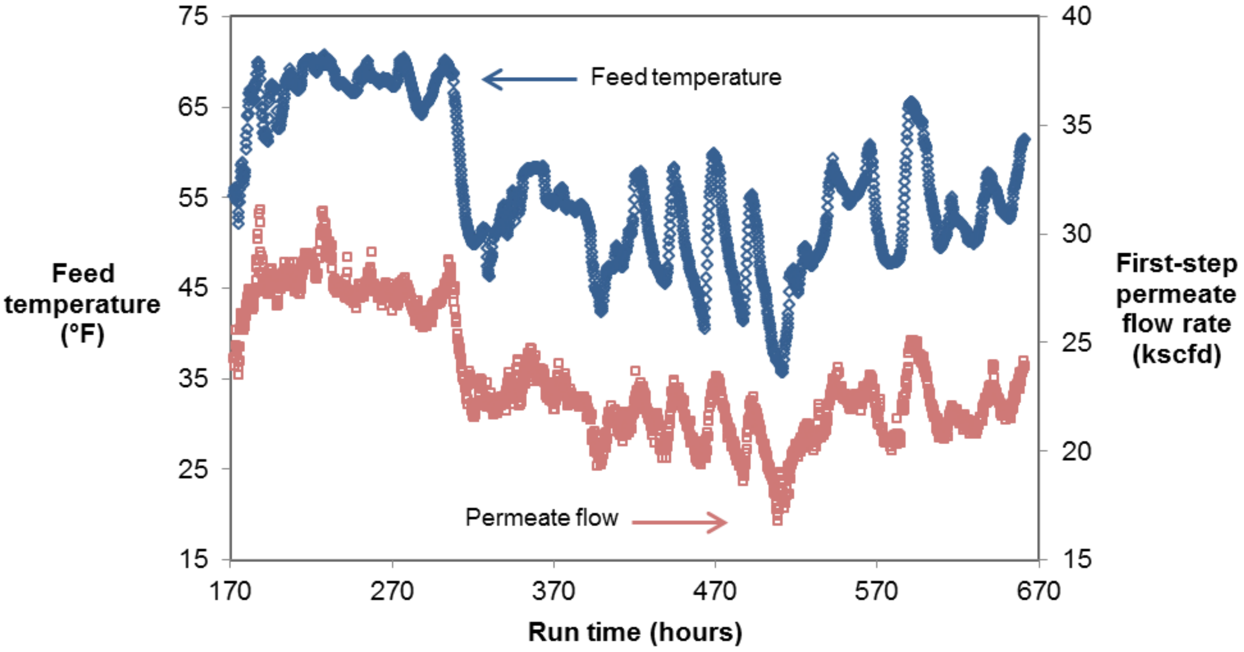


Figure 3.23. Temperature swings in the flue gas feed to the first step from January 8-28, 2013, showing impact on observed first-step permeate flow rate (kscfd).

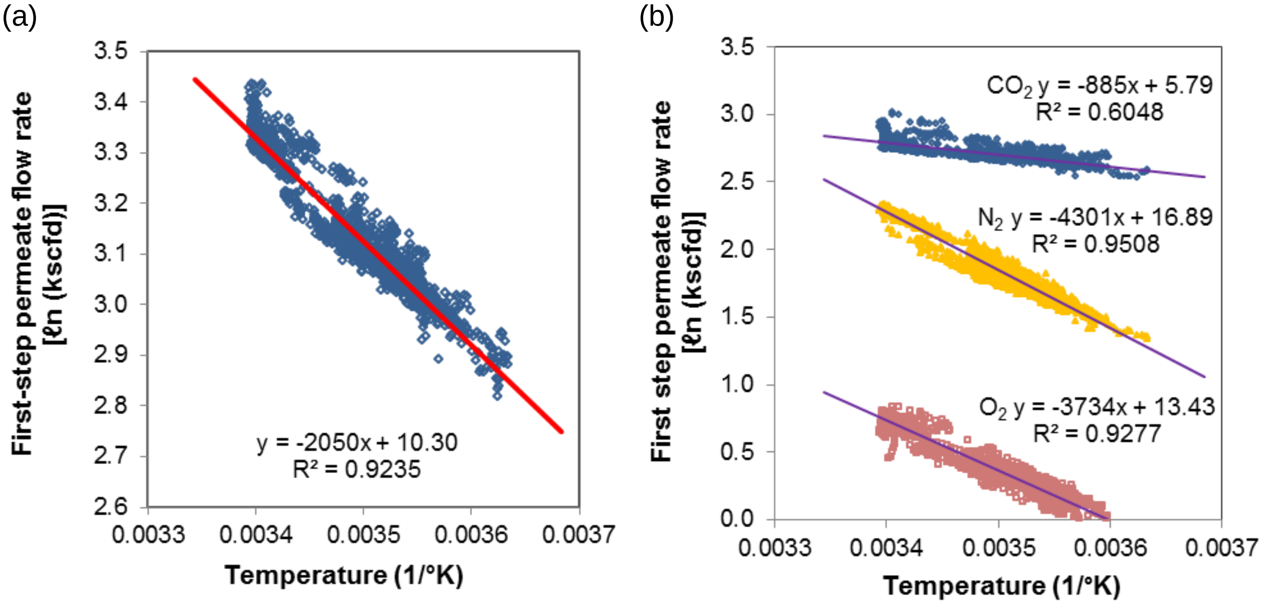


Figure 3.24. First-step permeate flow rate [ℓn (kscfd)] versus recorded temperature (1/°K) for data from January 8-28, 2013. (a) Combined gas flow; (b) individual gas flows.

The impact of temperature on the overall system performance is shown in Figure 3.25, which plots CO_2 purity in the first step permeate and overall carbon capture rate against system temperature. As temperature increases, the carbon capture rate increases because the CO_2 permeance of the modules is higher at higher temperature. On the other hand, the purity of the captured CO_2 decreases at higher temperatures, partly because the CO_2/N_2 selectivity of the modules decreases as temperature increases but also since higher stage-cut at higher temperature also lowers CO_2 purity. Considering the relatively large temperature range of 30 °F, the variations in capture rate are quite modest, even though the permeate CO_2 content ranged from 56 to 74%. This is another example of self-leveling performance, since almost doubling the permeate rate (from about 17 to 31 kscfd) only marginally increased the total carbon capture rate. In the future, a large capture system could compensate for temperature change by taking membrane area on or off-line to control product rate, adjusting the pressure ratio, or the system could be simply built and insulated for operation at constant temperature.

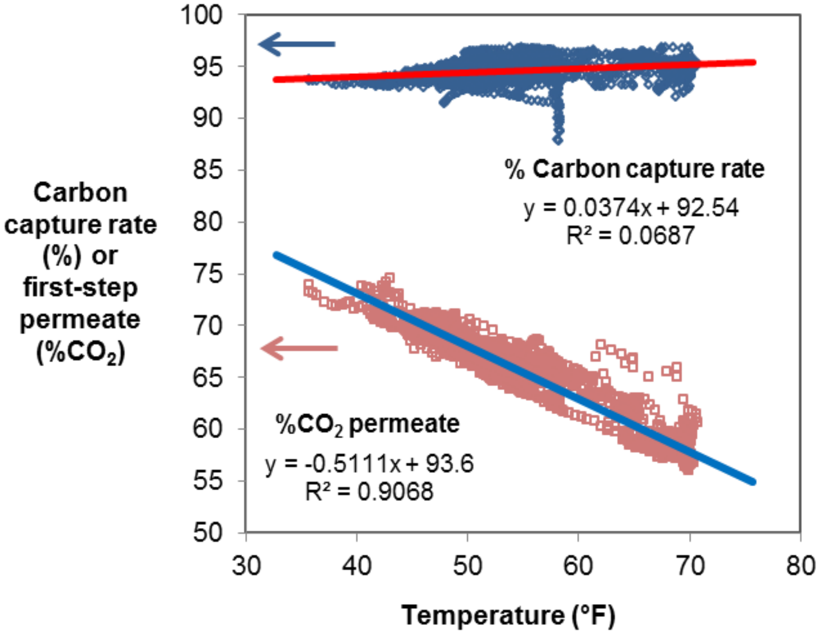


Figure 3.25. Carbon capture rate and $\%CO_2$ in first-step permeate as a function of system operating temperature for data from January 8-28, 2013.

During this test period in January 2013 the system showed relatively high carbon capture rates from 93-97%. These modules were constructed with membranes that had high selectivity for CO_2/N_2 which is further enhanced by operating at relatively cool January temperatures. These were the highest carbon capture rates observed over the years of this project.

3.8 Performance of Sweep Modules in Step Two

Figure 3.26 shows a good correlation between the CO_2 content of the residue streams from the first and second steps. Because the first step does about 50% CO_2 removal, the first step residue has half the CO_2 content of the feed. The data set separates into two regions depending upon whether all or half the modules are operated in the first-step. The band with the feed content of 5-7% CO_2 occurs when the first step is operated at half capacity, thereby leaving more CO_2 content in the first-step residue. The band with the feed content of 4-5.5% CO_2 occurs when all modules are running in the first-step. The fraction of CO_2 removed by the sweep modules is slightly higher when the first step operates with half capacity and when higher CO_2 concentrations are entering the sweep modules, which should generate higher driving force. This increase is reflected in Figure 3.26, which shows that when 50% of the modules are operating in step-one, the data is shifted slightly to the right of the "all modules" data.

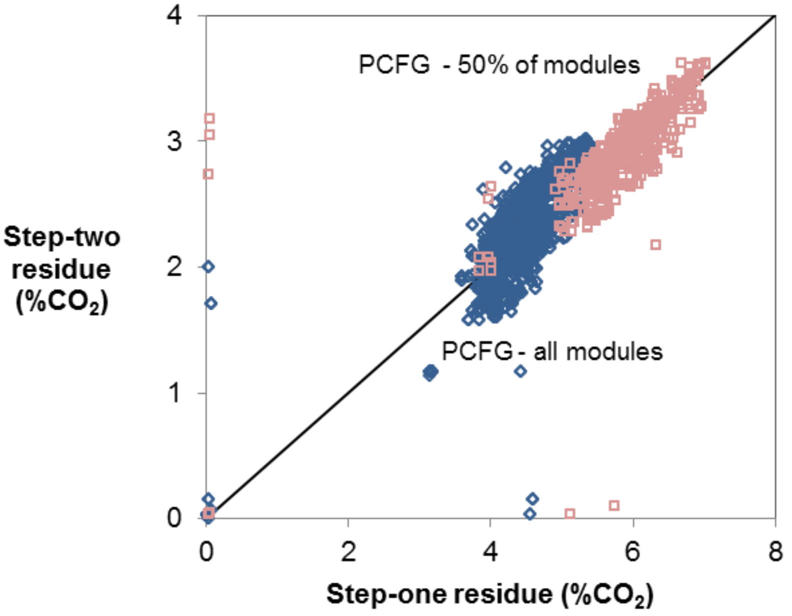
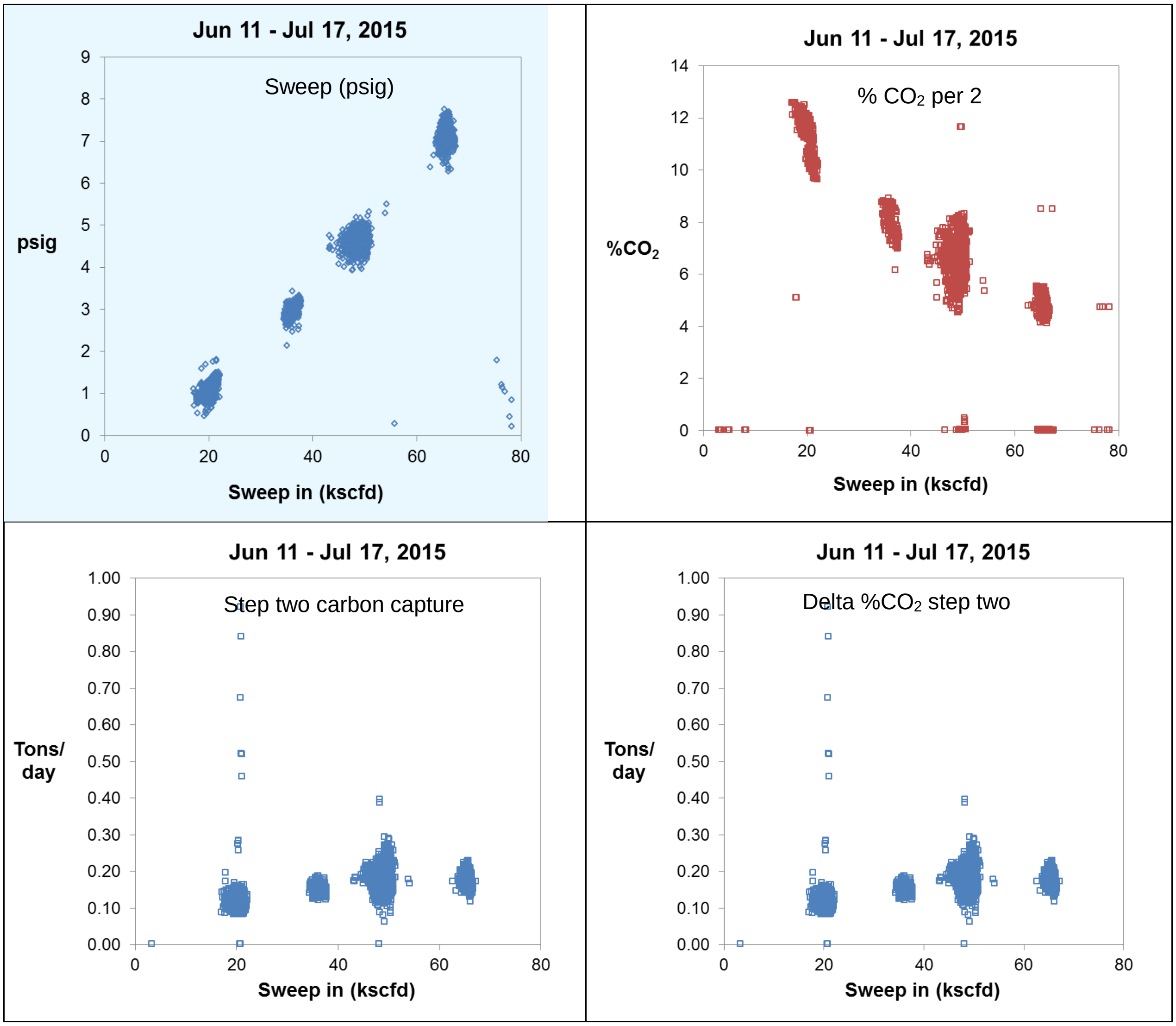


Figure 3.26. Data from January 23 to March 17, 2014, showing that the CO_2 content of the step-two residue is about half of the step-one residue. The number of modules (50% or all) refers to step one with cross-flow modules where the residue is the feed stream for step two. The number of sweep modules in step two remained constant. The diagonal line represents exactly 50% reduction in CO_2 concentration.

In 2015 the second step was operated with higher selectivity sweep modules when compared to the modules used in Figure 3.26. The advantages of a better CO_2/N_2 ratio is less permeation of nitrogen into the permeate, which will dilute both the CO_2 and O_2 concentrations of the sweep-out gas. Permeate flow rates through the sweep modules were reduced from 25-30 kscfd to 5-10 kscfd for the higher selectivity modules. With one-third or less permeate rate for N_2, there is less dilution of both CO_2 and O_2 content in the sweep-out.

Figure 3.27 shows that the pressure drop on the sweep side drops with lower sweep rates. The CO_2 content in the permeate (sweep-out) also increases at lower sweep rates, because there is less dilution of CO_2 by the air sweep. Higher sweep rates improve the quantity of CO_2 removed (expressed as tonne/day of CO_2 recovered in step two). If the difference between the residue CO_2 content from step one minus the step two residue is examined, then in accordance with improved tonne/day at higher sweep rate, the delta in residue CO_2 is also improved. Clearly, trade-offs in performance can be made with lower sweep rates to remove sufficient CO_2 with less pressure drop. Another option is designing for higher CO_2 recovery rates via higher sweep rates and a high pressure drop, but also with less dilution of O_2 in the sweep-out because more air is brought in by the high sweep rate.


Jun 11 - Jul 17, 2015

Jun 11 - Jul 17, 2015

Jun 11 - Jul 17, 2015

Jun 11 - Jul 17, 2015

Figure 3.27. Performance of step-two sweep modules at end of run in 2015 with variations in air sweep rate from June 11 to July 17, 2015.

3.9 Modeling Performance of Spiral-Wound Sweep Modules

Average values for the last month of operations (June 11-July 17, 2015) for the inlet feed gas temperature (80 °F), residue flow for step one (108 kscfd), residue pressure for step one (22 psig), and residue CO_2 content for step one (7.44%) were determined. These values set the initial boundary conditions for the sweep modules in step two and were used as input values for MTR's ChemCad process simulator to model ideal countercurrent performance. These calculated values are plotted against the observed field data in Figure 3.28.

Most of the variability in the field data was due to the broad range of feed gas temperature (from 70-102 °F). The data generated at 50 kscfd sweep rates averages 0.19 tonne/day CO_2 removal in step two at an average temperature of 83 °F. With the "no sweep" value at 0.031 tonne/day, this represents a 6.1 fold improved capture rate with these sweep modules. The ideal countercurrent sweep is calculated to be 0.27 tonne/day which places the sweep modules at 70% efficiency (100% * 0.19 / 0.27). For a sweep module without any requirement for vacuum equipment, this is an excellent rate of CO_2 removal.

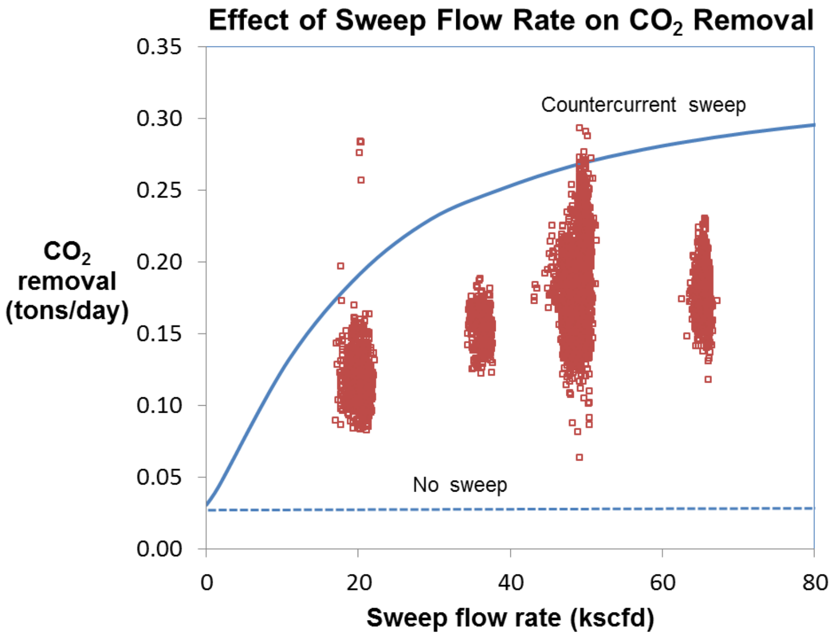


Figure 3.28. Calculated curve for perfect countercurrent sweep and the "no sweep" reference value compared to field data (squares) from June 11 to July 17, 2015.

3.10 Carbon Capture Rate and Enrichment Factor Steady Over Time

Figure 3.29 shows an expanded time frame for results reported in the period from March 1 to March 17, 2014 near the end of this campaign, with all modules running in the first step and providing stable operation. Looking at shorter times allows for more details to be seen. Over this period of almost 400 hours of uninterrupted operation, the feed was at 12% CO_2, the first-step permeate was generally greater than 50% CO_2, the first-step residue was less than 5% CO_2, and the second-step residue was less than 3% CO_2. These measured concentrations were used, together with pressures and flow rates, to calculate some of the important performance parameters for this application. Among these parameters are:

- the carbon capture rate, defined as the mass of CO_2 removed in the membrane permeates (first and second steps) divided by the mass of CO_2 in the feed gas, and
- the enrichment factor, defined as the ratio of CO_2 concentration in the first-step permeate to that in the feed.

As shown in Figure 3.29, the carbon capture rate ranges between 83-91% during this test period. This level of CO_2 removal is in the range of the DOE target for carbon capture from a pulverized coal power plant (90%). The CO_2 enrichment factor for step one ranges between 4 and 5 over the course of the test campaign. This value is important because ultimately the captured CO_2 will need to be purified and compressed for use in enhanced oil recovery or sequestration. The higher the membrane enrichment factor, the lower the cost of subsequent purification/compression. An enrichment factor of greater than 4 is consistent with our expectation for these operating conditions, and when combined with selective recycle to the boiler to bring the feed CO_2 concentration to near 20%, it would produce a CO_2-enriched permeate stream of sufficient purity to make final purification/compression cost effective. Both sets of data in Figure 3.29 are flat and stable over these 400 hours of operation with extension of the total run time now up to 1,300 hours.

Another performance value of importance to this program is the carbon capture rate, in tonne/day, which is shown as a function of time in Figure 3.30. The system was designed to treat a flue gas slipstream containing 1 TPD of CO_2 and capture up to 90% of this CO_2. This is precisely what happens at the beginning of the test period when 0.9 TPD was captured. The slight decline in the TPD of CO_2 captured over this time frame occurs because the CO_2 content in the flue gas feed stream dropped from 12.2% to 11.0%. This meant that progressively less CO_2 was being supplied to the membrane system, and therefore less mass of CO_2 was captured, although the percentage CO_2 capture rate remained constant at about 90%. Figure 3.30 also shows that for this particular arrangement of modules, the bulk of the carbon capture occurs in step one with the cross-flow modules. In step two, the sweep modules capture fewer tonnes of CO_2 because most of the CO_2 has already been removed from the gas fed to these modules.

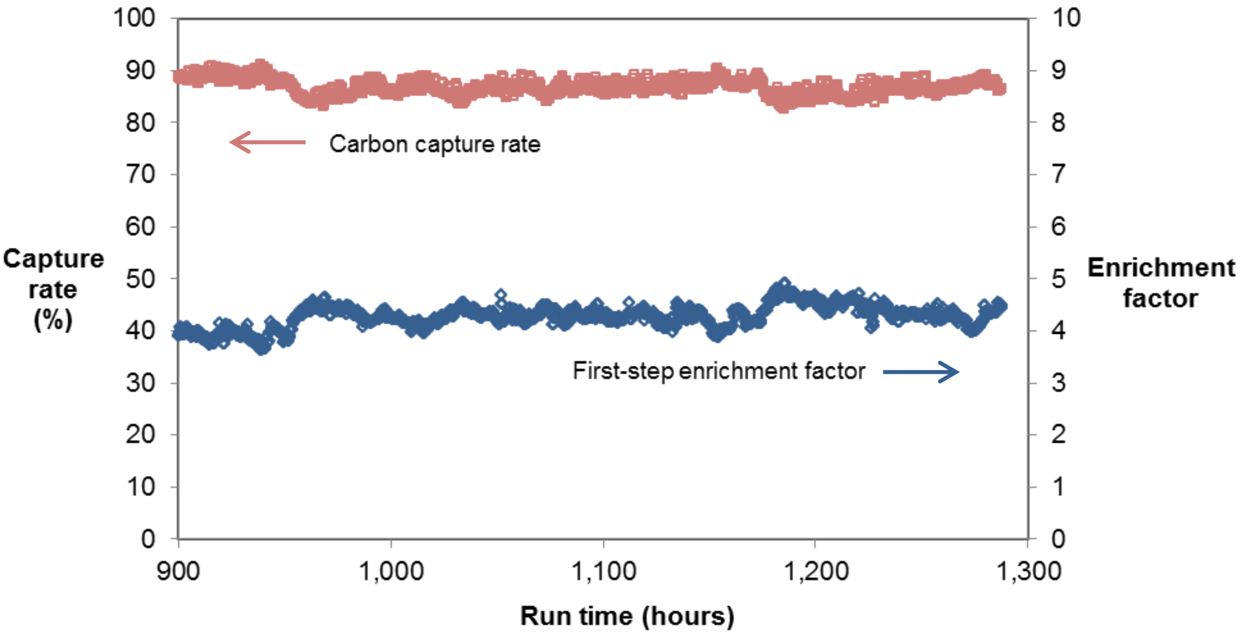


Figure 3.29. Expanded time scale from March 1-17, 2014, showing total carbon capture rate; and the enrichment factor for first-step CO_2 capture.

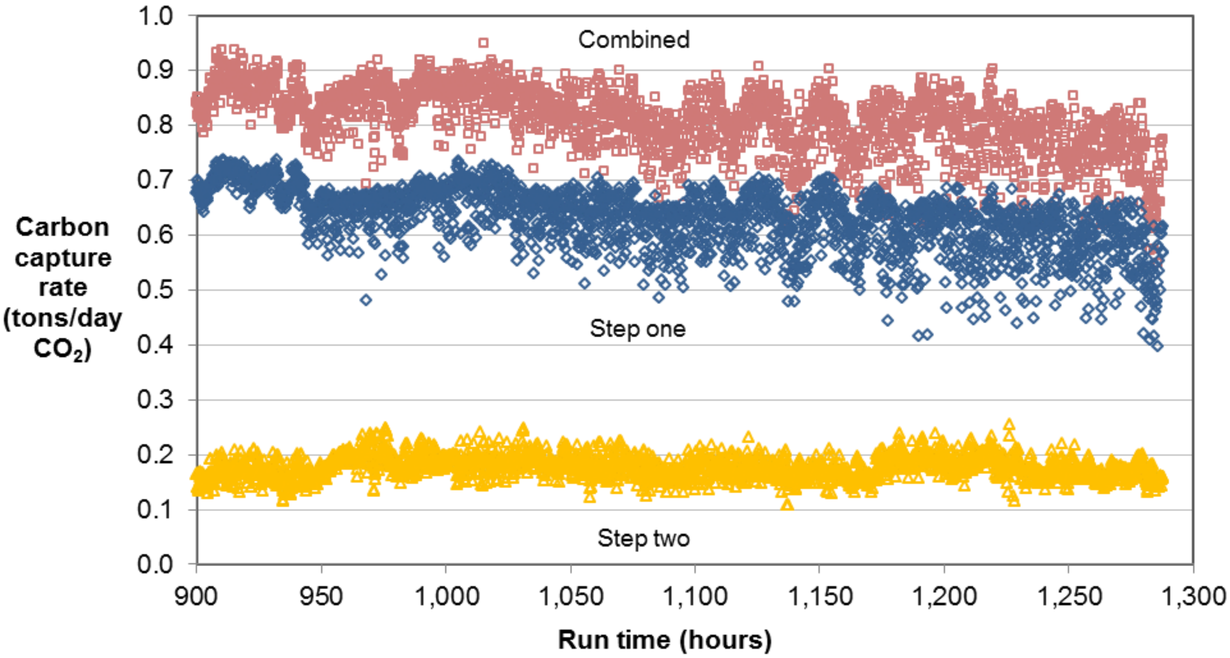


Figure 3.30. Expanded time scale from March 1-17, 2014, showing the tonnes/day of CO_2 captured from the step one modules, the step two modules, and the combined total. The slow decline in step one total carbon capture rate that was observed was because the feed fell from 12.2% to 11.0% CO_2 over this time period.

3.11 Chapter Summary and Conclusions

During 2012-2015, four extended test campaigns, each over 1,000 hours in length, showed that stable Polaris membrane module performance for carbon capture from real flue gas can be achieved. Full-scale spiral-wound cross-flow and sweep modules were developed for this project and were shown to operate effectively. Important for multi-year operations, the modules also demonstrated stable performance over multiple shutdowns and restarts of the test system.

Over three and half years of testing, the system achieved a total of 11,466 hours of operation with 9,100 hours on flue gas and 2,366 hours on air when flue gas was not available. Out of the 9,100 hours of testing on flue gas, 1,933 hours were on diluted flue gas during the natural gas simulation test.

Technical challenges were observed during the first year of trials that needed to be addressed. The dry screw compressor initially used to pressurize the feed gas suffered from corrosion problems and was finally replaced with a liquid ring compressor. It was also observed that the Polaris module performance was lost if the wet flue gas was allowed to condense onto the membrane as a strongly acidic water. Therefore, the modules were subsequently purged with air during shutdown periods. Once these changes were made in fall 2012, the system operated reliably over the remainder of its time at NCCC.

Carbon capture rates of better than 90% were achieved with coal-derived flue gas. The system also generated data for simulated flue gas derived from natural gas; although the unit was not designed for high recoveries from this alternative flue gas, capture rates of greater than 80% were observed.

While clear differences in the separation performance of the various module designs were observed, the impact on overall pilot plant performance was less than what might have been expected. This was due in part to mass balance requirements linking the CO_2 concentrations in the first-step permeate and residue streams, and the observation that high permeate flows weakened the permeate vacuum, and therefore lowered the available pressure ratio of the system. These competing factors made the system performance self-leveling in some respects and led to clearly defined bands of available performance ranges, depending on system feed and module set-up.

Significant responses to various operating conditions have been noted, particularly with respect to the choice of membrane and the number of operating modules, the system temperature, the available vacuum, and the stage-cut. The choice of module configuration sets clear boundaries on the performance envelope that can be expected. Another major influence is the system operating temperature since warmer temperatures increase permeation rates (with concurrent loss in selectivity), but this result is convoluted because higher permeation rates mean both larger stage-cuts and less available vacuum.

In 2015, the second step was operated with higher selectivity sweep modules. The advantages of a better CO_2/N_2 ratio is less permeation of nitrogen into the permeate. With one-third or less the permeate rate for N_2, there is less dilution of both CO_2 and O_2 content in the sweep-out stream.

Higher CO_2 and O_2 content in this stream are both beneficial for the proposed two-step membrane process for CO_2 recovery.

Lessons learned from this 1 TPD bench-scale system were applied to the design and construction of a 20 TPD small pilot plant, which will be the focus of Chapter 4.

4. DESIGN, FABRICATION, AND FIELD OPERATION OF A 20 TONNES PER DAY (20 TPD) CO_2 CAPTURE MEMBRANE SYSTEM

This chapter provides a summary of the design, construction and operation on a coal-fired flue gas slipstream of a 20 TPD MTR membrane CO_2 capture small pilot system. Scale-up of the MTR Polaris capture system from the bench-scale (1 TPD) unit described in the previous chapter to this 20 TPD small pilot system was the primary objective of this project. During the first budget period, it was decided that the small pilot would be tested at the NCCC for direct comparison with other advanced post-combustion capture technologies being evaluated at this site.

4.1 Design and Construction

The design work for the 20 TPD small pilot system included performance feedback and lessons learned from the 1 TPD CO_2 capture membrane system that had been in operation at NCCC since spring of 2012. The 20 TPD system was a two-level design with membrane modules located on the upper level and all rotating and associated equipment on the lower level. A general arrangement drawing of the skid design is shown in Figure 4.1.

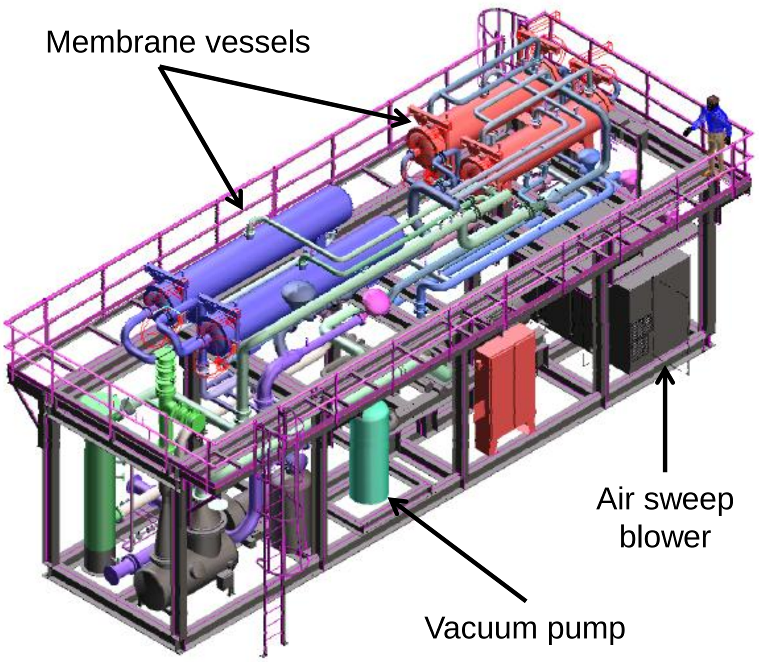


Figure 4.1. General arrangement drawing of the MTR 20 TPD CO_2 capture membrane system.

The 20 TPD small pilot system was designed to take a slipstream from an existing 880 MW_e supercritical pulverized coal unit at Plant Gaston in Wilsonville, AL. The design of the small pilot system does not include recycle of CO_2 captured in the sweep step back to the boiler as it would have an inconsequential impact on the boiler and the coal-fired flue gas CO_2 concentration.

For this system, the flue gas enters the first-step spiral-wound membrane modules where a vacuum on the permeate side creates a driving force for CO_2 to permeate the membrane. The flue gas then enters the second step of the system where air is swept on the permeate side to create a driving force for further depleting the flue gas of CO_2. In this process the flue gas exiting the second step, CO_2-laden air from the permeate side of the second step, and the high CO_2 concentration permeate stream from the first step are all sent to the stack. This slipstream design allows for both steps of the MTR process design (vacuum permeate first step and air-sweep module for the second step) to be tested at the 20 TPD scale on real coal-fired flue gas. A simplified process flow diagram of the 20 TPD system is shown in Figure 4.2.

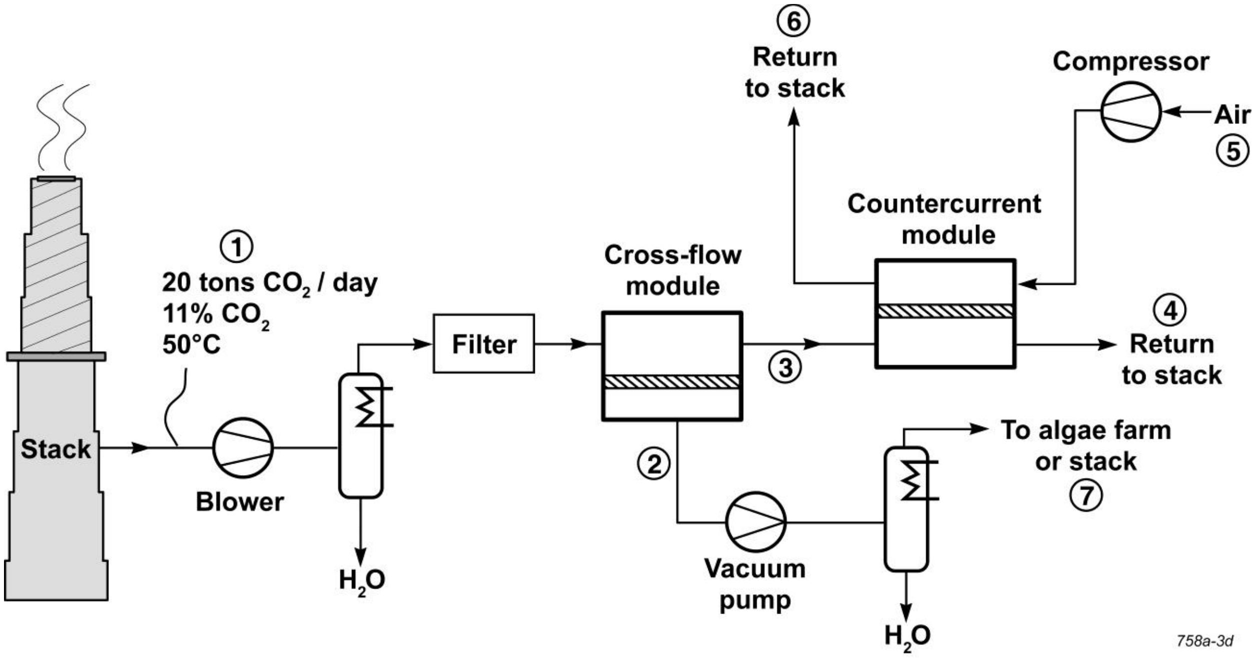


Figure 4.2. Process flow diagram of the 20 TPD small pilot system.

The 20 TPD unit design included cost reduction features such as commercial-sized, bundled modules within multi-tube housings, and low-cost plastic end caps and tubing (described in Chapter 2). The bundled modules containing multi-tube housings also reduce piping complexity of the system and limit the number of valves and other piping components needed for operation. In addition to cost savings, the bundled modules reduce the overall size and footprint of the CO_2 capture system, which is an important consideration for a full scale system (500 MW_e). The bundled module vessels were fabricated by Johansing Iron Works (Oakland, CA) and individual membrane modules were loaded into the vessels prior to shipment to the 20 TPD system fabrication shop. Figures 4.3 and 4.4 show details of the bundled module vessels before shipment for final assembly.

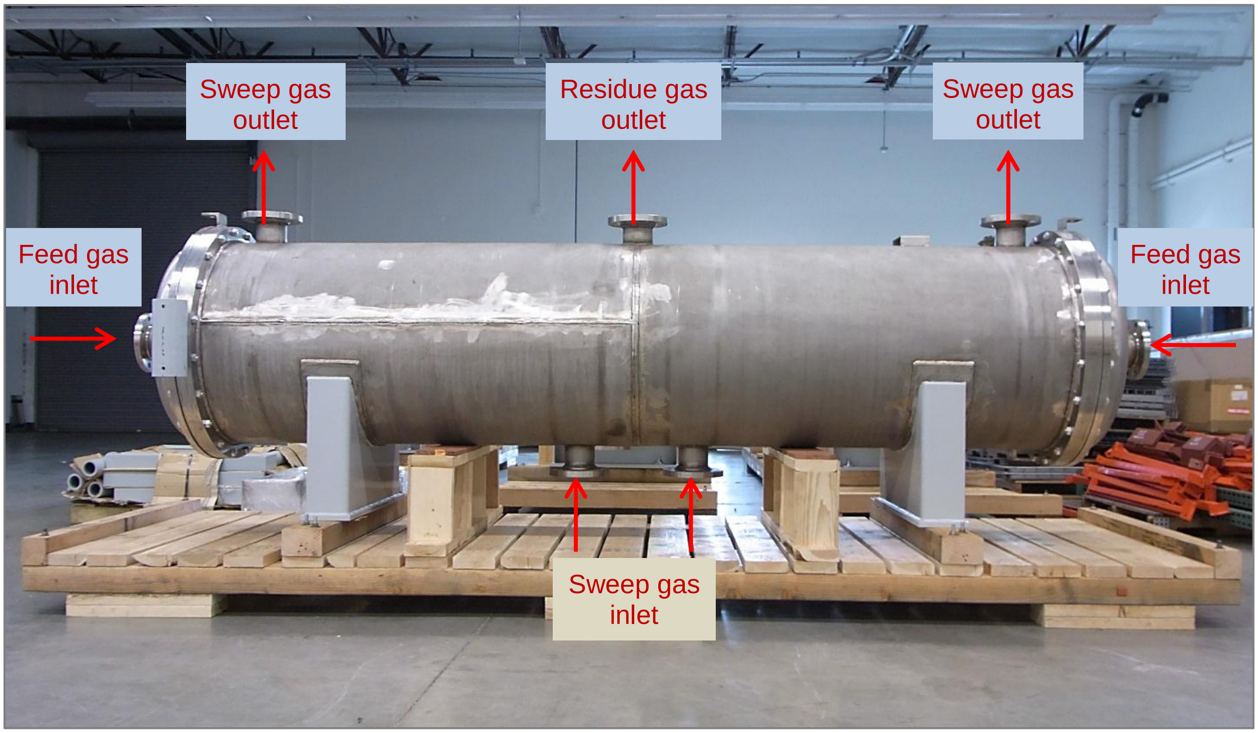


Figure 4.3. Photo of one of the completed sweep multi-insert pressure vessels used on the 20 TPD system.

Figure 4.4. Cross-sectional view of a bundled module pressure vessel with spiral-wound modules installed.

Glex, Inc. in Houston, TX was selected as the fabricator for the 20 TPD system. The scope of work at the Glex site included skid fabrication and integration of all system components, including flue gas feed compressor, permeate vacuum pump, gas/liquid separator, bundled membrane pressure vessels, and all skid instrumentation. Figures 4.5 and 4.6 show both levels of the 20 TPD system during fabrication at the Glex facility.

Figure 4.5. Lower level of the 20 TPD system containing rotating equipment and associated vessels during fabrication.

Figure 4.6. Upper level of the 20 TPD system containing membrane module vessels and associated piping during fabrication.

4.2. Installation and Commissioning

The NCCC in Wilsonville, AL was chosen as the host site for the 20 TPD system for testing on a coal-fired flue gas slipstream. As part of the NCCC, the Post-Combustion Carbon Capture Center (PC4) is a slipstream test facility utilizing flue gas from Alabama Power Plant Gaston Unit 5, which is an 880 MW_e supercritical pulverized coal unit. NCCC started site preparation for the 20 TPD system in April 2014, which kicked off the installation work. The installation was completed in mid-July 2014, and MTR and NCCC started the initial system shakedown and commissioning afterwards. Work continued until late August, when a planned three-month flue gas outage started at the site. Figures 4.7, 4.8, and 4.9 show pictures detailing the installation of the 20 TPD system at NCCC, as well as illustrating the overall size and footprint of different separation systems at NCCC designed to remove similar amounts of CO_2 from coal-fired flue gas streams. Figures 4.7 and 4.8 illustrate the simple modular nature of membrane systems where all fabrication can be completed at a fabrication shop prior to shipping to site. Once unloaded at the site, only the connecting piping between various skids and electrical hook-ups need to be taken care of before shakedown, commissioning, and operation can commence. As there is no on-site construction required for the membrane skids, downtime at the plant is minimized. In Figure 4.9, the large 150 ft. superstructure to the far right houses the NCCC Pilot Solvent Test Unit (PSTU), which tests various solvents for CO_2 capture on a scale of 10 TPD. In the center of Figure 4.8, the even larger structure is for an advanced solvent CO_2 capture system at a 20 TPD scale. In the far left of the figure is the MTR membrane CO_2 capture system also sized to process 20 TPD. The size and simplicity advantages of the membrane system are readily apparent.

Figure 4.7. Lower level of the 20 TPD system arriving at NCCC for installation.

Figure 4.8. A crane lowers the upper level of the 20 TPD system into place during installation.

Figure 4.9. Footprint comparison of equivalently-sized CO_2 capture separation technologies at NCCC.

Prior to the scheduled flue gas outage in fall 2014, the 20 TPD system was run on both air and flue gas for a short duration. The system operated as expected, with the exception that one of the bearings on the Nash liquid ring feed compressor ran hot. During the three-month outage, the feed compressor was sent to a Nash facility for inspection, installation of new bearings, and a test run to verify normal performance. The Nash compressor was then shipped back to NCCC and re-installed in the 20 TPD system in the fourth quarter of 2014. In early January 2015, six MTR engineers completed a three day operator training session for the 20 TPD system at NCCC. These engineers rotated to the NCCC site over the duration of the two test campaigns (January through June 2015) to support operation of the system. The nomenclature that NCCC uses for post-combustion test campaigns are PO-X, where the X indicates the run number. MTR operated the 20 TPD small pilot test system during PO-2 (January through late March 2015) and for a portion of PO-3 (early May through the end of June 2015) before decommissioning and removing the system in July 2015. While on-site, the MTR engineers were responsible for analyzing field data, troubleshooting any skid issues, and communicating with both NCCC and MTR personnel. Table 4.1 shows the tasks and their completion dates for the 20 TPD system installation and commissioning.

Table 4.1. Installation and Commissioning Tasks for the 20 TPD System at NCCC.

Task	Start	Finish	Performed by
Site pre-commissioning	04/30/14	05/27/14	NCCC
Heat tracing connections	05/21/14	06/17/14	NCCC
MCC construction	05/12/14	05/23/14	NCCC
120 VAC power termination	05/26/14	06/17/14	NCCC
Scrubber punch list items	05/28/14	06/05/14	NCCC
Compressor punch list items	05/27/14	06/05/14	NCCC
Vacuum pump punch list items	05/27/14	06/03/14	NCCC
Sweep gas blower punch list items	05/27/14	06/20/14	NCCC
Loop check and test sweep gas controls	06/23/14	06/24/14	MTR
Membrane skid punch list items	06/02/14	06/23/14	NCCC
Atlas Copco commissioning	06/23/14	06/25/14	Atlas Copco
Loop check and test compressor and vacuum pump controls	06/23/14	06/24/14	MTR
Nash commissioning	06/27/14	12/01/14	Nash
Inlet scrubber commissioning	06/10/14	06/11/14	MTR
Loop check and test membrane controls	06/13/14	06/16/14	MTR
PLC testing	05/21/14	07/18/14	MTR
Sequence check	07/01/14	07/18/14	MTR
SAT test	07/04/14	07/18/14	NCCC and MTR
Commission entire system	07/07/14	01/18/15	NCCC and MTR
Operator training	08/04/14	01/14/15	NCCC and MTR
Installation of plate-and-frame sweep skid	02/16/15	03/02/15	NCCC and MTR
Commissioning of plate-and-frame skid	04/13/15	04/17/15	NCCC and MTR
20 TPD system operation	01/18/15	06/30/15	NCCC and MTR

Various shakedown and commissioning issues were identified and fixed while running the system on air only in January 2015. Once flue gas became available, the system was commissioned on January 18 during PO-2. Skid operations were fine-tuned while the complete system was running until the belts on the vacuum pump failed on January 24. The system continued to run without the first-stage vacuum pump until January 28, when the system was shut down to install new belts on the vacuum pump and other miscellaneous tasks, including installation of a small pump on one of the gas cabinets by NCCC to ensure the gas analyzers have the proper pressure to accurately measure different gas stream compositions.

4.3 System Operation

The system ran intermittently on flue gas in February 2015 with shut downs due to cold weather issues on the NCCC side (pre-scrubber and cooling water), and miscellaneous system issues on the MTR side. When operating, the system consistently had an overall CO_2 capture rate of ~86%. The typical CO_2 removal performance of the 20 TPD system over a week in February is shown in Figure 4.10. As illustrated, the daily performance fluctuations are minimal; ambient temperature has less of an influence on performance for the 20 TPD system than it does for the smaller 1 TPD system. Possible reasons for this include better heat tracing and insulation of the skid to insulate the system from ambient conditions and temperature control of the flue gas to the first-step modules on the 20 TPD system. In addition to steady state data, parametric tests were also conducted with different combinations of the first and second step module vessels at reduced flow rates to determine the performance of membrane in each vessel.

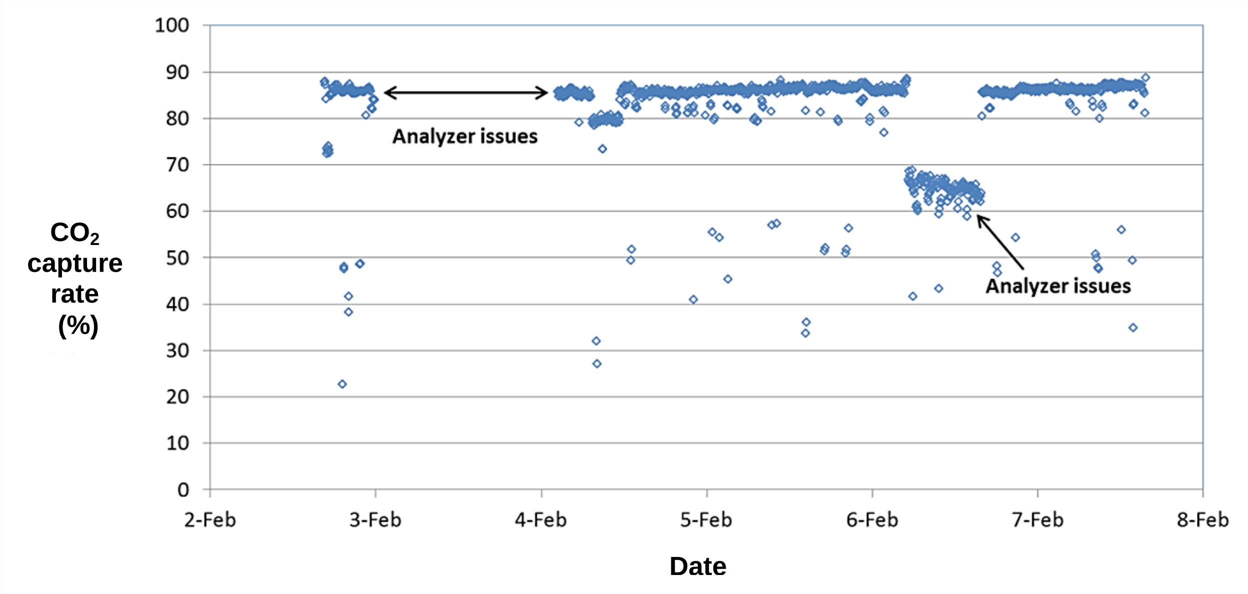


Figure 4.10. Overall CO_2 capture rate of the 20 TPD system during PO-2 steady state operations.

On February 15, the 20 TPD skid was shut down for the planned installation and integration of the plate-and-frame membrane module sweep skid and a small water sampling skid. Developed under a separate DOE project (DE-FE0007553), the plate-and-frame sweep module skid is a novel type of membrane module specifically designed for the high-gas-flow, low-pressure sweep operation required for affordable membrane-based CO_2 capture at coal power plants. The main driving force for the development of the novel module design was to reduce the sweep-side pressure drop, compared to that measured on spiral-wound modules. In addition, the plate-and-frame sweep modules are constructed out of plastic components and packaged to contain 20 to 25 times the membrane area of traditional spiral-wound modules used for CO_2 capture, which reduces the cost and complexity of manifolding the membrane modules together, as well as the footprint of the complete membrane system. The combination of energy savings, reduced module and skid costs, and reduction in the overall membrane system footprint make this technology attractive for future scale-up. The 500 m^2 plate-and-frame sweep module skid was sized to be a direct replacement to the spiral-wound sweep modules installed during fabrication of the 20 TPD system and is shown in Figures 4.11 and 4.12.

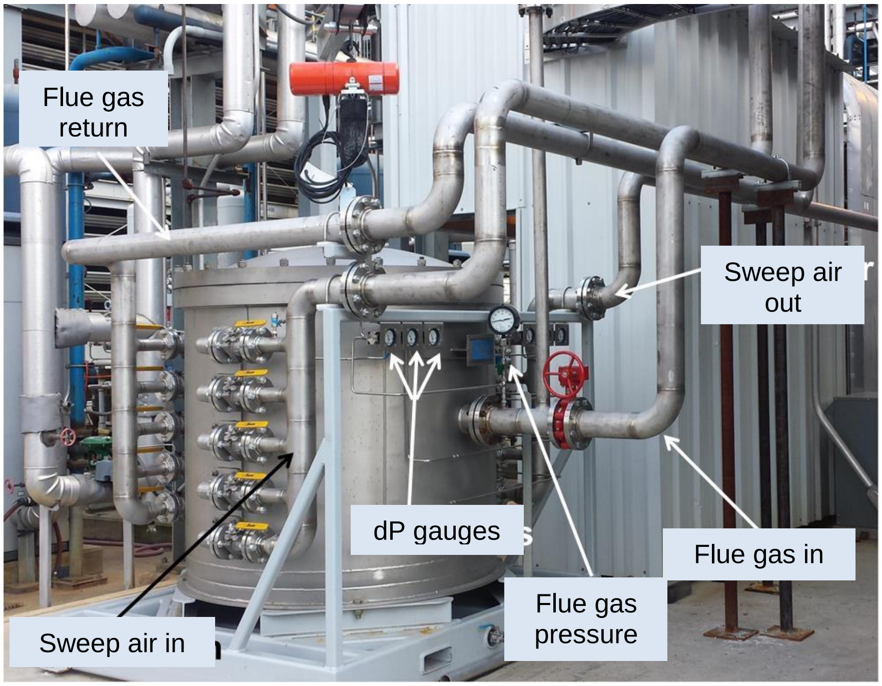


Figure 4.11. Plate-and-frame membrane module sweep skid after installation at NCCC.

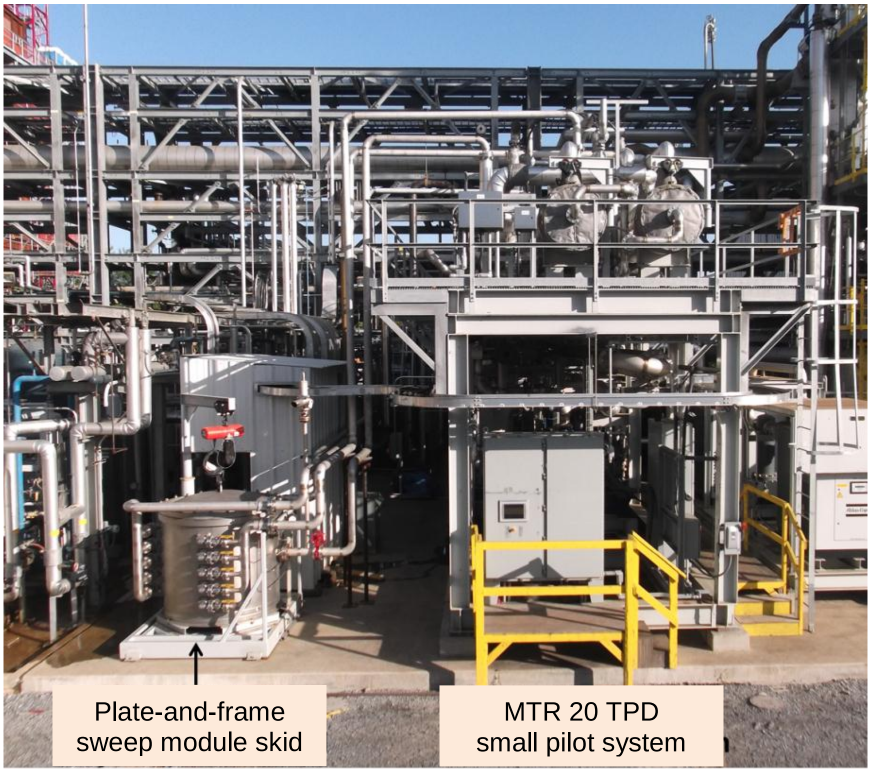


Figure 4.12. Comparison of the plate-and-frame membrane module sweep skid next to the larger 20 TPD small pilot unit at NCCC.

The 20 TPD system was restarted using all spiral-wound modules (including sweep spirals on the second step) on February 28, and ran steadily until cold weather caused cooling water supply issues on the NCCC side, which tripped the system on March 6. Due to this system trip, a critical flow switch on the 20 TPD system was damaged. Because of the lead time to replace this part and the approaching end date of the campaign, the 20 TPD system did not operate for the short remainder of PO-2. The cumulative run time of the system on flue gas during the PO-2 campaign was approximately 400 hours, with CO_2 capture rates ranging from 85% to 95%. The CO_2 capture rate of the 20 TPD system during the PO-2 campaign is shown in Figure 4.13.

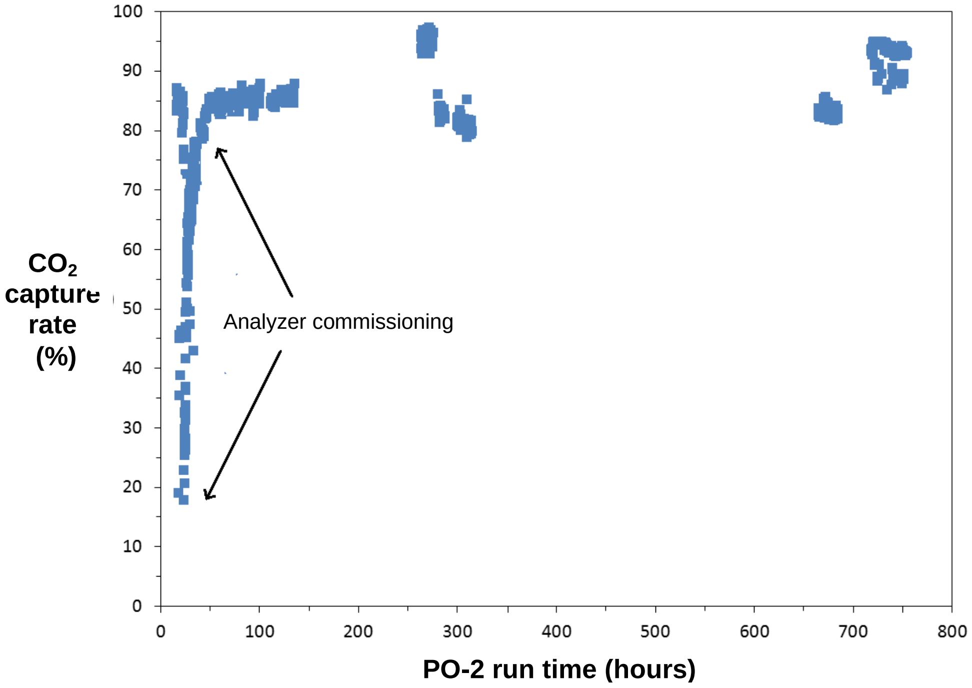


Figure 4.13. CO_2 capture rate of the 20 TPD system during the PO-2 post combustion flue gas campaign at NCCC.

In April, while flue gas was unavailable between NCCC campaigns PO-2 and 3, MTR commissioned the plate-and-frame membrane module sweep skid with air. As previously mentioned, this skid tests an advanced sweep module developed in a separate DOE program. Based on laboratory tests, this plate-and-frame module should have significantly lower sweep-side pressure drop, compared to spiral-wound sweep modules tested during the PO-2 campaign.

Figure 4.14 summarizes the results of commissioning tests with air conducted in April. Figure 4.14(a) shows the feed-side pressure drop through the plate-and-frame module as a function of air flow rate on the feed-side of the module. The data points in this figure represent measurements from the field at NCCC, while the solid line shows the expected performance based on laboratory tests. As expected, increased flow rate increases the pressure drop in nearly a linear manner (laminar flow). The measured field data is slightly below the lab data, although after some flow rate calibration adjustments, the agreement between the data sets appears even better. In all cases, the feed-side pressure drop is <1 psi, which is lower than the values used in our systems analysis (0.1 bar or 1.5 psi). Figure 4.14(b) shows the comparison between field and lab data for sweep-side pressure drop as a function of air sweep flow rate. Once again, the agreement between field and lab data is quite good, and most importantly, the pressure drops are low (<1 psi). The sweep-side pressure drop is where we believe the new plate-and-frame module has a significant advantage over traditional spiral-wound sweep modules. Previous test results on the 20 TPD system showed sweep-side side pressure drops of >4 psi for spiral-wound modules, so a key aspect of plate-and-frame module testing is to confirm that they have lower pressure drop when operating with flue gas.

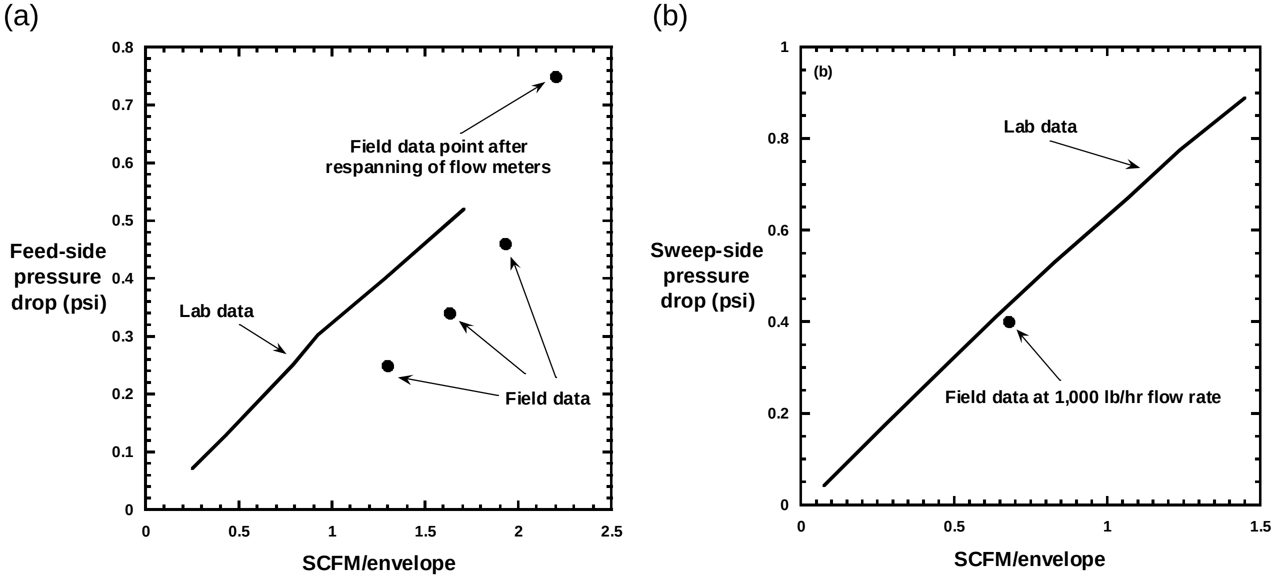


Figure 4.14. Comparison of sweep-side pressure drop through the plate-and-frame modules in field commissioning tests at NCCC and in the MTR lab.

Flue gas was introduced to the 20 TPD system on May 8 for the start of the PO-3 post-combustion campaign. The main goal of the PO-3 campaign was to operate the plate-and-frame sweep skid as the second step of the 20 TPD system for a minimum of 500 hours to validate the lab performance values and demonstrate the stability of the new module design under real coal-fired flue gas conditions. Another goal was to determine the various operating parameters for the 20 TPD system under summer conditions (ambient temperatures of 90 °F and higher), compared to the sub-freezing winter conditions of PO-2.

Downtime for the 20 TPD system was minimized during PO-3, due to lessons learned during PO-2. Figure 4.15 shows that the 20 TPD system with the plate-and-frame sweep skid consistently captured CO_2 at a rate of 85% or higher throughout the test campaign. The system was online with flue gas more than 70% of the time between May 8 and June 30, with the majority of downtime due to flue gas outages outside of our control. NCCC did strive to make flue gas available as much as possible, and even provided flue gas to the system through a different blower when the blower and pre-scrubber to Bay 3 (the location of the 20 TPD system) were down for repairs.

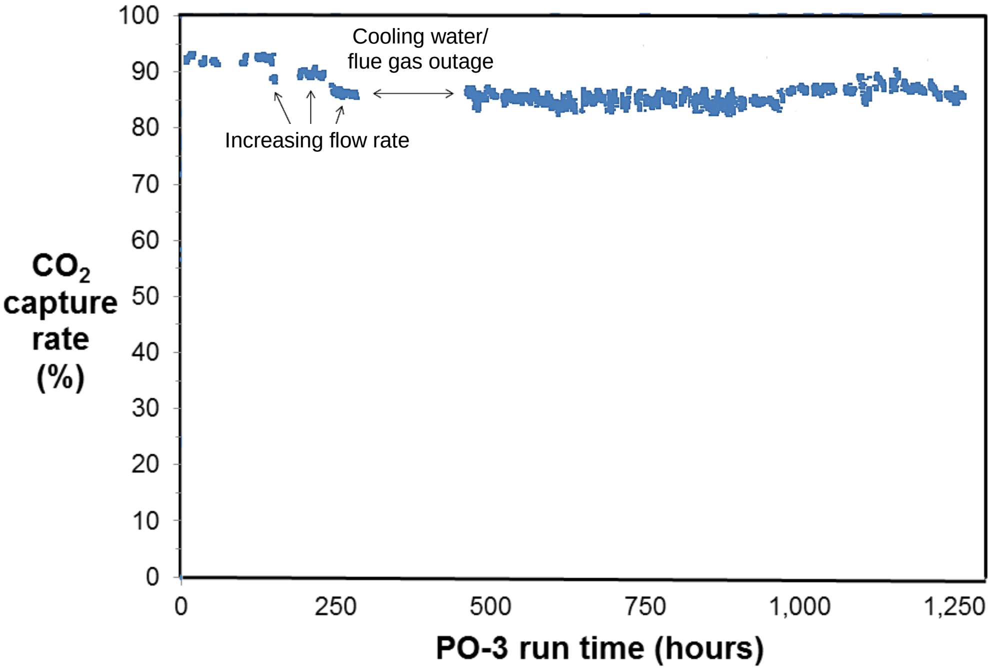


Figure 4.15. CO_2 capture rate of the 20 TPD system during the PO-3 post combustion flue gas campaign at NCCC.

Figure 4.16 shows an example of the CO_2 enrichment produced by the MTR membrane system by tracking the CO_2 content of various gas streams over the period of June 10 – 13. As shown in the figure, the typical flue gas feed to the membrane systems was 11-12% CO_2. The first step permeate was enriched about 4 times to 45% CO_2, while the treated gas leaving the residue stream contained about 2% CO_2. This performance is very similar to that achieved by the smaller 1 TPD system and is in agreement with our design calculations. For an integrated system (as opposed to the slipstream testing at NCCC), the CO_2 recycled to the boiler would increase the feed CO_2 content to around 20% (instead of 11% at NCCC). As a result, with an expected permeate enrichment of 3 to 4 times, this permeate stream would contain >70% CO_2, which would then be sent to a liquefaction system for purification and pumping to sequestration pressures.

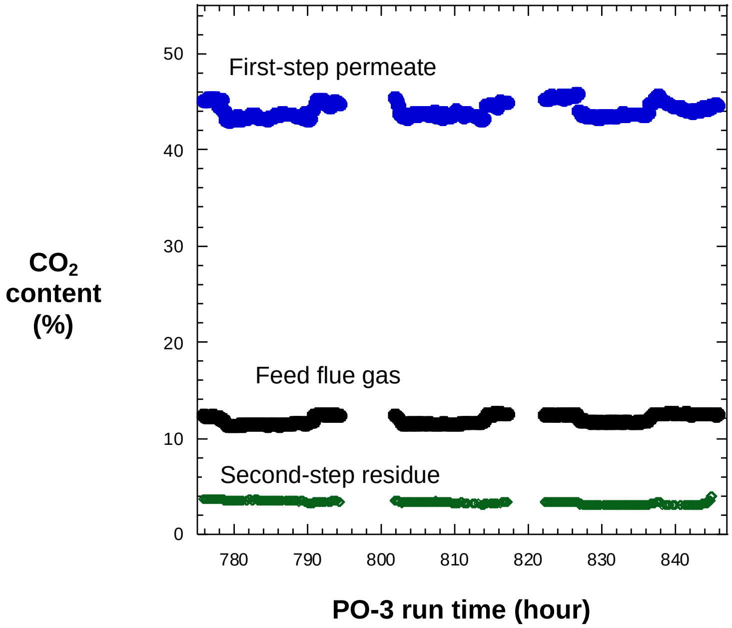


Figure 4.16. CO_2 content of the feed and permeate gas streams of the first step of the 20 TPD system, and the residue of the second step during operation from June 10 – 13, 2015.

The plate-and-frame sweep skid was on-line as the second step of the 20 TPD system for the duration of PO-3. This allowed for direct comparison to the spiral-sweep module performance used as the second step during PO-2, as well as parametric testing. Figure 4.17 compares the sweep-side pressure drop for the two types of modules under the same conditions at NCCC. The field data for the plate-and-frame modules is consistent with lab data and confirms the significantly lower pressure drop with the new module design. The plate-and-frame sweep module has roughly four times lower pressure drop, compared to the spiral-wound sweep modules tested during PO-2. For a full scale power plant (550 MW_e), this reduction in pressure drop would amount to an energy savings of ~10 MW_e.

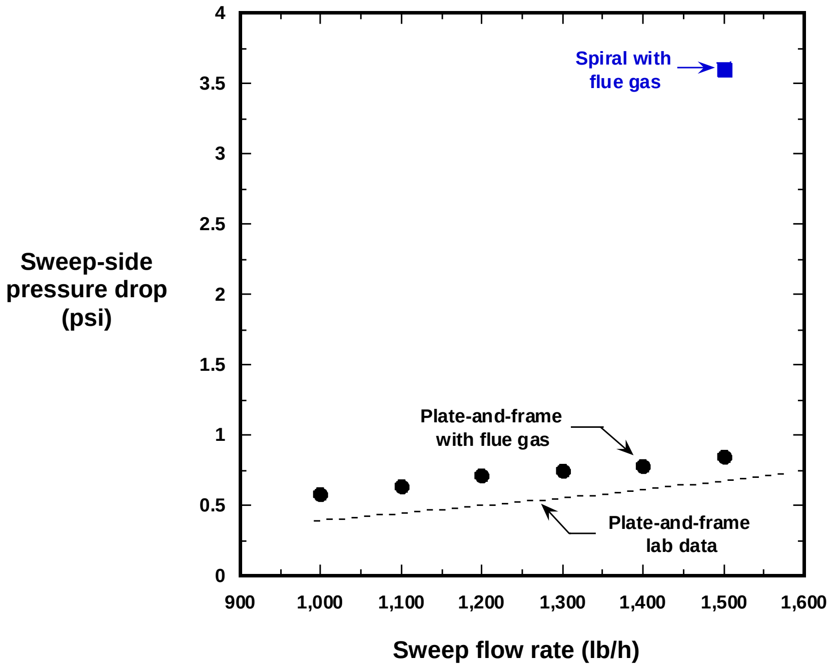


Figure 4.17. Sweep-side pressure drop comparison of plate-and-frame and spiral-wound sweep modules during testing with flue gas on the 1 MW_e system at NCCC.

The performance of the plate-and-frame sweep module skid was sensitive to operating parameters. Figure 4.18 details the influence of the flue gas feed pressure on the CO_2 removal rate of the plate-and-frame sweep skid. During these tests, the sweep air flow rate to the permeate side of the plate-and-frame module skid was maintained at 1,500 lb/hr. The solid line in the figure shows the expected CO_2 removal by the skid without air sweep. The difference between the expected capture without sweep and the experimentally measured CO_2 removal performance with sweep underscores the importance of sweep in pressure-ratio limited applications, such as CO_2 removal from coal-fired flue gas. As the flue gas feed pressure to the plate-and-frame sweep skid increases, the net outcome is a higher CO_2 partial pressure driving force, which results in a higher CO_2 removal rate for both cases with and without an air sweep. Over the feed pressure range tested (27.5 – 37.5 psia), the CO_2 removal rate was improved by over 40%, by increasing the flue gas feed pressure to the plate-and-frame sweep module skid.

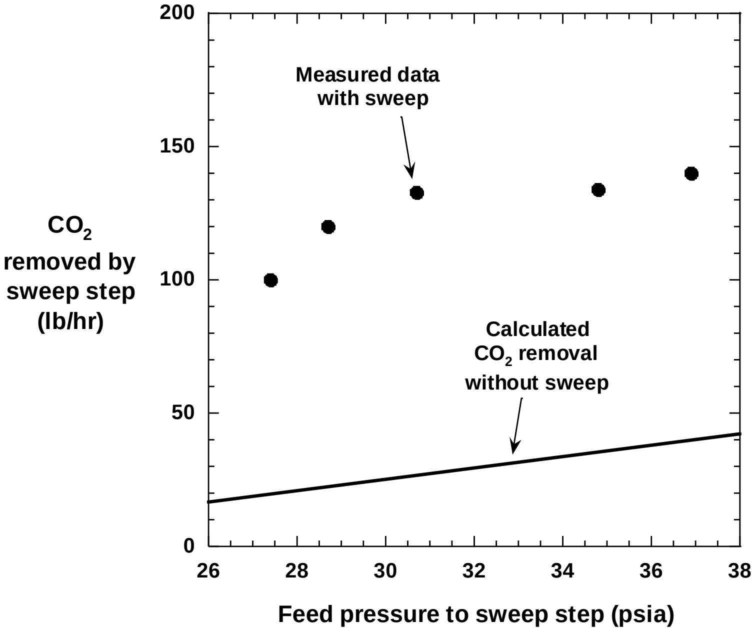


Figure 4.18. Influence of the flue gas feed pressure on the CO_2 removed by the plate-and-frame sweep module skid during the PO-3 campaign at NCCC.

Another important adjustable parameter for the second step is the sweep air flow rate to membrane modules. Figure 4.19 shows the influence of the air sweep flow rate on the CO_2 removed by the plate-and-frame sweep module skid at a constant flue gas feed pressure of 28 psia. As in Figure 4.18, a solid line represents the expected CO_2 removal by the skid without air sweep. In this case, with a constant flue gas feed pressure and no sweep, the CO_2 partial pressure driving force is constant, so the expected CO_2 removal without sweep is the same for all comparable sweep conditions. Increasing the sweep air flow rate to the membrane modules allows for greater dilution of CO_2 on the permeate side, resulting in higher CO_2 partial pressure driving force and removal rates. Over the relatively small range of sweep air flow rates tested, the CO_2 removal rate was increased by ~45%.

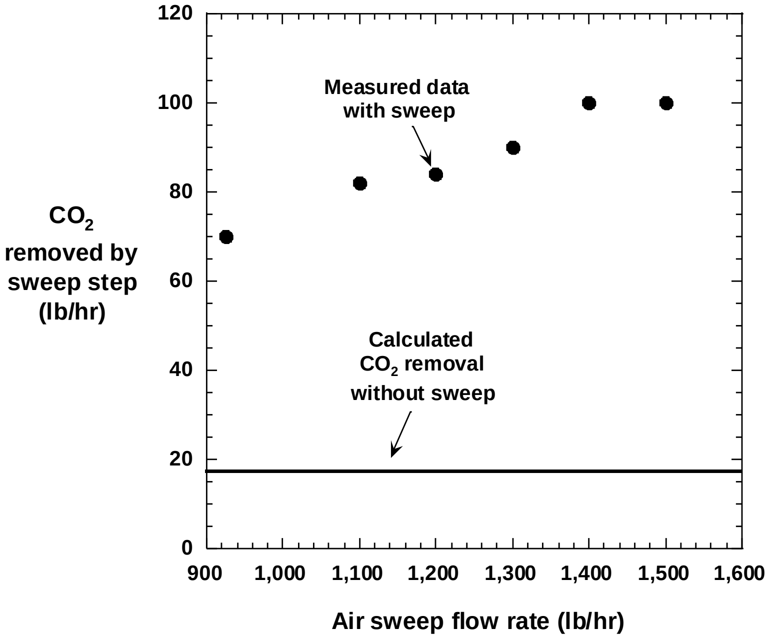


Figure 4.19. Influence of the sweep air flow rate on the CO_2 removed by the plate-and-frame sweep module skid during the PO-3 campaign at NCCC.

During PO-3 operation, the 20 TPD system ran on flue gas for ~1,000 hours with stable CO_2 capture rates and minimal downtime. This campaign exceeded the goal of obtaining 500 hours of operation for the plate-and-frame sweep module, and demonstrated the superior pressure drop performance of this module.

The 20 TPD system was shut down on June 30 by an MTR engineer, as previously scheduled, to begin the decommissioning and removal process of the system from the NCCC site. A long-term shut down, including draining of all process and cooling water from the skids, was conducted by MTR, while the remaining decommissioning and removal tasks were handled by NCCC personnel. All skids associated with the 1 MW_e system were removed by the end of July and transported to Ohio for future integrated operation of the system with recycle of CO_2 to an appropriately-sized boiler at B&W research facilities under DE-FE0026414.

4.4 Chapter Summary

Overall, the 20 TPD system was successfully installed, commissioned, and operated during two post-combustion campaigns at NCCC. This small pilot test system highlighted the simple modular nature of membrane systems as all skid fabrication took place prior to arrival at NCCC. The NCCC field test was also the first field test where bundled modules containing multi-tube housings were utilized for cost reduction and the plate-and-frame sweep module skid developed specifically for low pressure, low pressure drop sweep applications was used. Other test highlights include:

- Stable system operation meeting design specifications under both sub-freezing winter conditions and high temperature, high humidity summer conditions;
- 400 hours of operation during PO-2 with CO_2 capture rates ranging from 85% to 95% with spirals as the sweep step;
- 1,000 hours of stable operation during PO-3 achieving CO_2 capture rates over 85% with the plate-and-frame module design as the sweep step;
- About 4 times lower sweep-side pressure drop for plate-and-frame sweep modules, compared to spiral-sweep modules operated under the same field test conditions.

With these achievements, all of the success criteria associated with the 20 TPD system design, construction, and operation set at the beginning of the project were met. The MTR membrane technology is now ready for future scale-up to large pilot-scale. Concurrently, as mentioned above, the 20 TPD unit has been removed from NCCC and will be reused for integrated boiler testing at B&W in a separate follow-on project to demonstrate the selective CO_2 recycle process.

5. IMPACT OF RECYCLED CO_2 ON BOILER PERFORMANCE

5.1 Introduction

One of the key differences between conventional post-combustion CO_2 capture approaches and the MTR membrane capture process is the selective recycle of CO_2 to the boiler used by the MTR process. As described earlier, the selective recycle step improves the efficiency of capture by increasing the concentration of CO_2 in the flue gas and reducing the fractional CO_2 removal required by the capture step in a single pass. However, unlike other end-of-the-pipe capture approaches, the MTR process changes operating conditions in the boiler. By recycling CO_2, the boiler combustion air is diluted so that the concentration of oxygen is reduced compared to conventional operation. An important question is what affect these changes, if any, have on the boiler performance. Quantifying the impact of recycled CO_2 on boiler performance, and its effect on the overall competitiveness of the MTR capture process was one of the objectives of this project.

Our partner, B&W, conducted two studies during this project to better understand the feasibility of recycling CO_2 to a coal boiler. The scope and primary results from these studies were as follows:

1. In budget period 1, a computational fluid dynamics (CFD) analysis was performed using B&W's proprietary COMO^{SM} software to estimate flame stability, gas compositions, and heat distributions for various amounts of CO_2 recycle in two common pulverized coal (PC) boilers. This analysis suggested it was feasible to operate PC boilers with recycled CO_2, while having a relatively modest impact on performance. Detailed results from this study are provided in the B&W topical report titled "Effect of Synthetic Secondary Air on Boiler Performance", dated March 15, 2012. The key recommendation from this report was to validate the modeling results with pilot scale testing.

2. In budget period 2, a pilot-scale test of boiler operation with CO_2-laden combustion air was conducted on B&W's SBS-II 0.6 MW_e research boiler. This test demonstrated flame stability with varying CO_2 recycle levels and quantified the impact of the CO_2 level on the overall boiler efficiency. Detailed results are provided in the B&W topical report titled "Effect of CO_2-Enriched Air on Combustion Performance: Pilot-Scale Evaluation" dated December 17, 2013. The key recommendation from this report was to test the full MTR capture process integrated with a coal boiler.

In this chapter, the most important findings from these B&W studies are summarized. The estimated impact of recycled CO_2 on boiler efficiency determined from these studies was incorporated into the final TEA of the MTR capture process described in Chapter 6.

5.2 Background

Figure 5.1 shows a diagram of the MTR two-step CO_2 capture process integrated with a pulverized coal-fired power plant. Secondary air going to the boiler is first routed to a membrane sweep module and used to remove CO_2 from flue gas. Dilution of the secondary air by CO_2 causes the oxygen concentration of the combustion air going to the boiler to decrease. Primary air is not used in the membrane capture process and is routed to the boiler in the conventional manner.

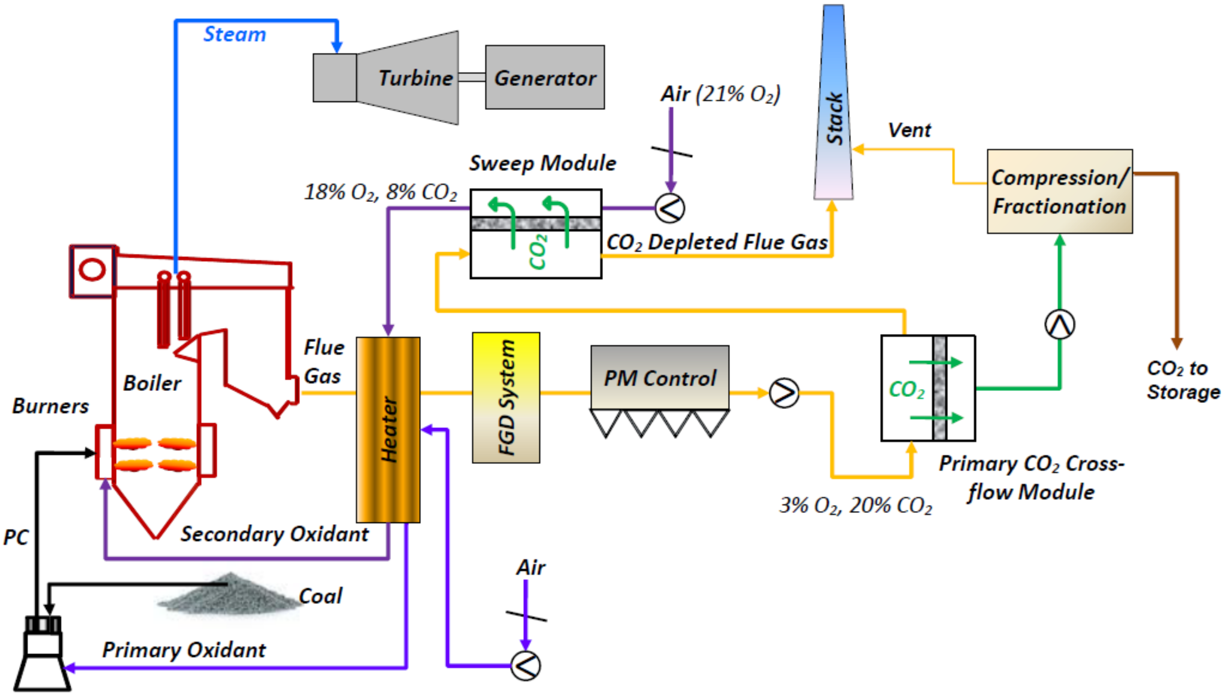


Figure 5.1. Simplified diagram of the MTR selective recycle CO_2 capture process integrated with a coal-fired power plant.

For the CFD modeling study, the effect of recycled CO_2 was evaluated on two typical opposed-wall pulverized coal-fired boilers: one firing bituminous coal and the other firing Powder Basin River (PRB) coal. The first boiler was a Carolina-type radiant boiler (RB) firing bituminous coal, with a nominal rating of 600 MW. The second boiler was a spiral wound ultra-supercritical universal pressure (SWUP) boiler firing PRB coal, with a nominal generating capacity of 690 MW. B&W's proprietary COMO^{SM} modeling software was used to evaluate the performance of the selected boilers and to examine the effect of changes in the secondary oxidant composition on flame stability. Performance was evaluated by comparing gas temperature, carbon monoxide and oxygen concentration distributions, carbon burnout, unburned carbon in fly ash and heat absorption on furnace walls.

Two sets of cases were modeled for each boiler. For each set, the amount of CO_2 recycled was varied to produce CO_2-enriched secondary air at reduced oxygen levels (20% to 17.5% by volume). In the first set, the overall boiler stoichiometry was held constant while varying amounts of CO_2 were recycled. To maintain a constant stoichiometry (or constant O_2 mass flow) while recycling CO_2 requires the secondary air flowrate to increase for these cases compared to normal air firing. In a second set of cases, the secondary air flowrate was held constant, but stoichiometry was allowed to vary. For these cases, as additional CO_2 is recycled, it replaces some of the secondary air, and therefore, the oxygen mass flow going to the boiler is reduced. From a capture standpoint, the constant secondary air flowrate case would be preferable because this approach generates a slightly smaller and more concentrated flue gas stream. However, it was expected that the constant flowrate case would have more of a negative impact on boiler performance than the constant stoichiometry case, so all previous base-case process designs and TEAs have assumed constant stoichiometry is required. The point of these different sets of modeling cases was to confirm whether this was true.

For the pilot testing in budget period 2, B&W's Small Boiler Simulator (SBS-II) was utilized. This facility consists of a multitude of components including a coal process area housing coal feeders and the mill, a fan room, the burner, a furnace with convection pass, heat exchangers and environmental control devices such as baghouses, scrubbers and SCR. A B&W DRB-4Z^{TM} low NO_x burner was used for this study. Two coals – a western sub-bituminous PRB and an eastern bituminous – were tested to study the effect of CO_2 addition to combustion air on boiler performance. To mimic CO_2 recycle by MTR's membrane process, CO_2 from a storage tank was added to the secondary air stream at SBS-II during experiments. Figure 5.2 illustrates the SBS-II equipment arrangement for testing including the addition of CO_2 to the secondary air line to simulate CO_2-enriched air operation. This figure also shows the site locations at which various process data were measured during testing. Preparation of the SBS-II for the CO_2-enriched air testing involved design and installation of the CO_2 supply system.

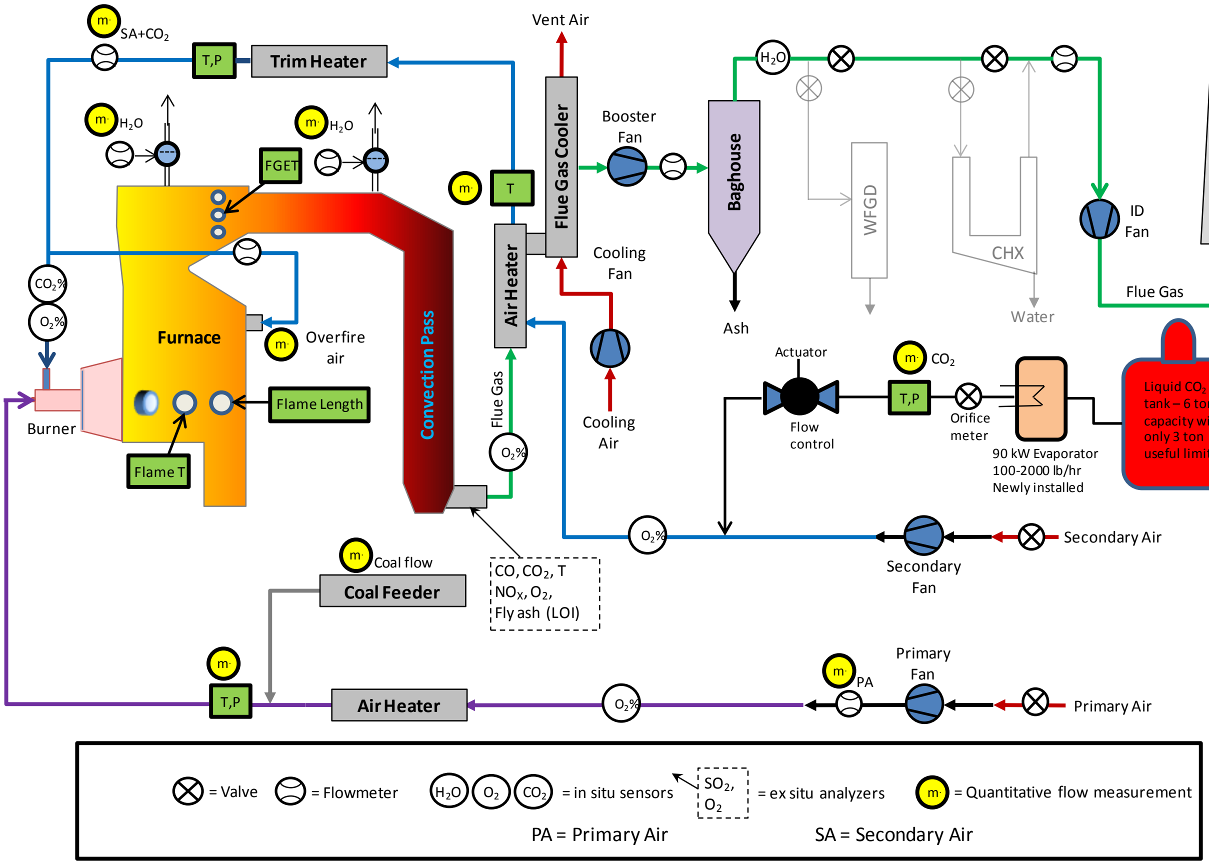


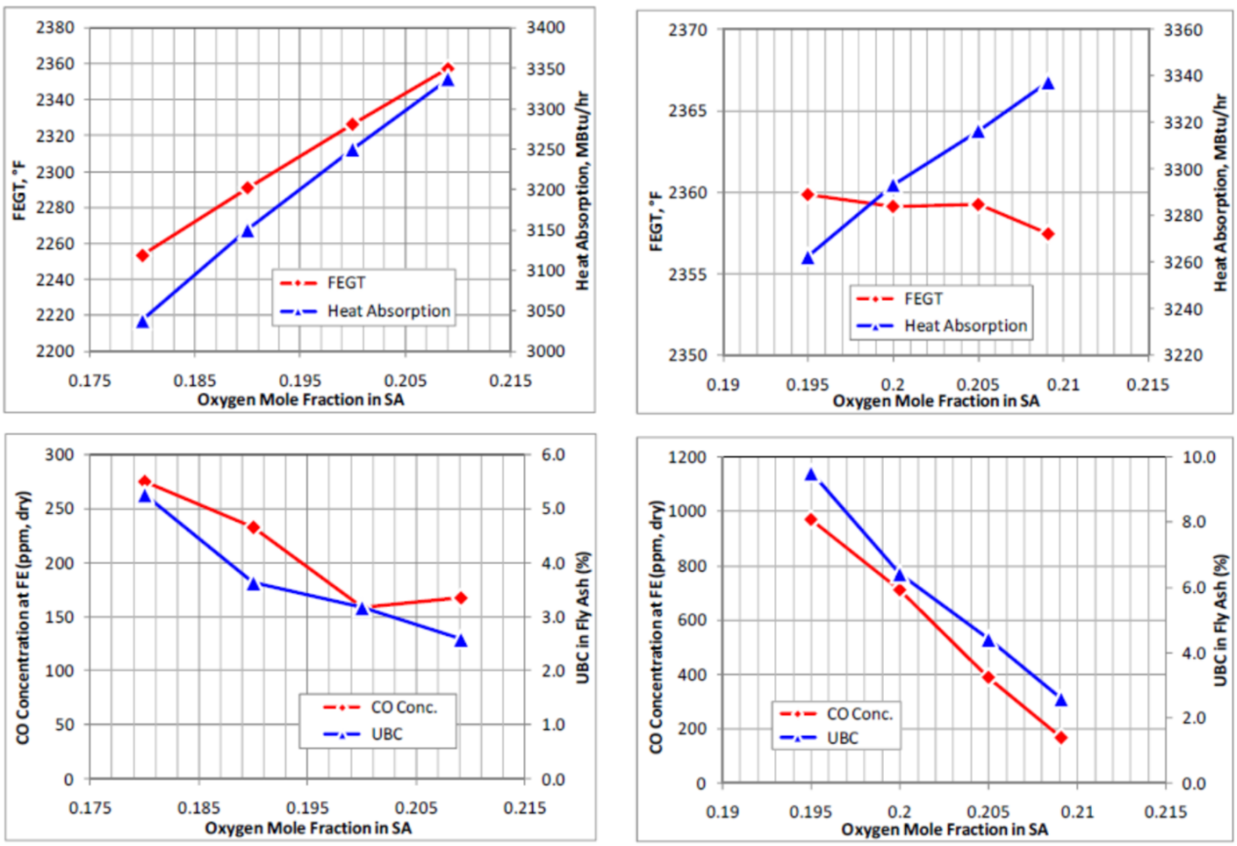
Figure 5.2. Schematic of the SBS-II equipment arrangement during testing with CO_2-laden secondary air.

5.3 Results of CFD Modeling Study

The key results from the B&W CFD modeling study can be summarized as follows:

- For all CO_2 recycle cases (corresponding to secondary air oxygen contents of 20.5% to 17.5% by volume), the models showed stable and attached flames at the burners.
- When the secondary air is diluted with CO_2 and the boiler stoichiometric ratio remains unchanged, both the boilers (RB and SWUP), firing a bituminous coal and a PRB coal, respectively, perform well in terms of combustion efficiency and the furnace exit gas temperature (FEGT).
- However, when the secondary air flow rate remains unchanged and the stoichiometry changes (oxygen mass flow rate in the secondary air decreases), CO concentration at the furnace exit and the unburned carbon (UBC) in fly ash become unacceptably high for both boiler types.

Figure 5.3 shows an example of these results for the SWUP boiler firing PRB coal. The x-axis in these plots shows the content of oxygen in the secondary air. As CO_2 recycle increases, the oxygen content decreases, moving from right to left on the plots. Figure 5.3(a) presents modeling results for the cases where boiler stoichiometry is fixed (secondary air flow increases as CO_2 recycle increase), while Figure 5.3(b) shows the same results for the cases where secondary air flow is fixed (variable stoichiometry). Note that unburned carbon and CO concentrations are much higher for the variable stoichiometry cases.


(a)

(b)

Figure 5.3. The effect of secondary air (SA) oxygen content on FEGT, heat absorption, CO concentration, and UBC for CO_2 recycle cases with (a) fixed stoichiometry and (b) variable stoichiometry.

Based on these modeling findings, fixed stoichiometry (variable secondary air flow rate) is required for retrofit of the MTR CO_2 capture process with recycle to the boiler. With fixed stoichiometry, the boiler steam performance was analyzed for each of the two boiler types at ~18% oxygen in the secondary air (corresponding to the amount of CO_2 recycle in the MTR base case design). This analysis produced the following results:

- For the series back-end, bituminous coal-fired boiler (RB), the increased flue gas flow rate caused by CO_2 recycle leads to about a 6% increase in heat absorption in the superheater and reheater tube banks. If the arrangement of the original bank heating surface is used, the spray flow of the superheaters and reheaters will increase. To minimize spray flow and maintain plant heat rate, some superheater and reheater surface may be removed.

- For the parallel back-end, PRB coal-fired boiler (SWUP), the higher flue gas flow rate due to CO_2 recycle also results in approximately a 6% increase in heat absorption in the superheater tube banks. If the arrangement of the original bank heating surface is used, the gas biasing dampers in the convection pass will not be able to divert enough flue gas away from the horizontal reheaters to limit the reheater heat absorption. One solution would be to remove some reheater heating surface or/and increase reheater spray.

In summary, the CFD modeling results suggest that CO_2 recycle in secondary air is feasible as a retrofit for either of the boiler configurations examined provided the boiler stoichiometry is maintained by increasing the secondary air flow. Under these conditions, the predicted combustion and heat transfer characteristics for CO_2-enriched air operation showed only a modest change from the baseline air-firing results. Based upon these favorable outcomes, a pilot-scale coal combustion and emissions performance evaluation with CO_2-enriched air was proposed as the next step.

5.4 Results of the Pilot Boiler Tests

The main objectives of the pilot boiler tests were to evaluate the effect of CO_2-enriched air on coal combustion and emissions performance and to ensure that CO_2 recycle does not cause any negative effects on boiler operation and reliability. Specifically, the testing addressed the impact of CO_2-enriched secondary air on following items:

- Flame stability, length and shape.
- Unburned combustibles in fly ash.
- Emissions, including NO_x and CO.
- Furnace exit gas temperature (FEGT).
- Radiant furnace and convection pass heat transfer and boiler thermal efficiency.

5.4.1. Flame Stability

Changes in flame stability and attachment from diluting the combustion air with CO_2 were monitored through visual observations. Necessary adjustments to the burner swirl vane angles and transition zone damper were made to achieve stable flames and reasonable pressure drop across the windbox. Optimum burner settings under air-firing operation were also used for CO_2-enriched testing for direct comparison of the performance results.

Based on visual observations, CO_2-enriched flames were brighter, more flared and more stable than flames witnessed under air-fired conditions. Flame attachment to the burner throat varied mainly with coal rank and to a lesser extent boiler stoichiometric ratio – but the effect of CO_2 addition clearly improved flame stability and quality under most test conditions. Conclusions from visual observation were well supported by data from the FlameDoctor, a B&W proprietary product that can assess flame characteristics. As shown in Figure 5.4, Kurtosis, skewness and dimension are all statistical variables that measure deviation from linearity (or a Gaussian distribution). A well-defined flame is expected to produce a nearly Gaussian signal distribution, and a detached/pulsing flame is expected to deviate from a Gaussian distribution. In essence, a low statistical parameter indicates a better flame. From these figures, it is clear that increasing CO_2 addition improves the stability of the flame considerably. This conclusion was derived from FlameDoctor data recorded on two different days (05/21/2013 and 06/18/2013).

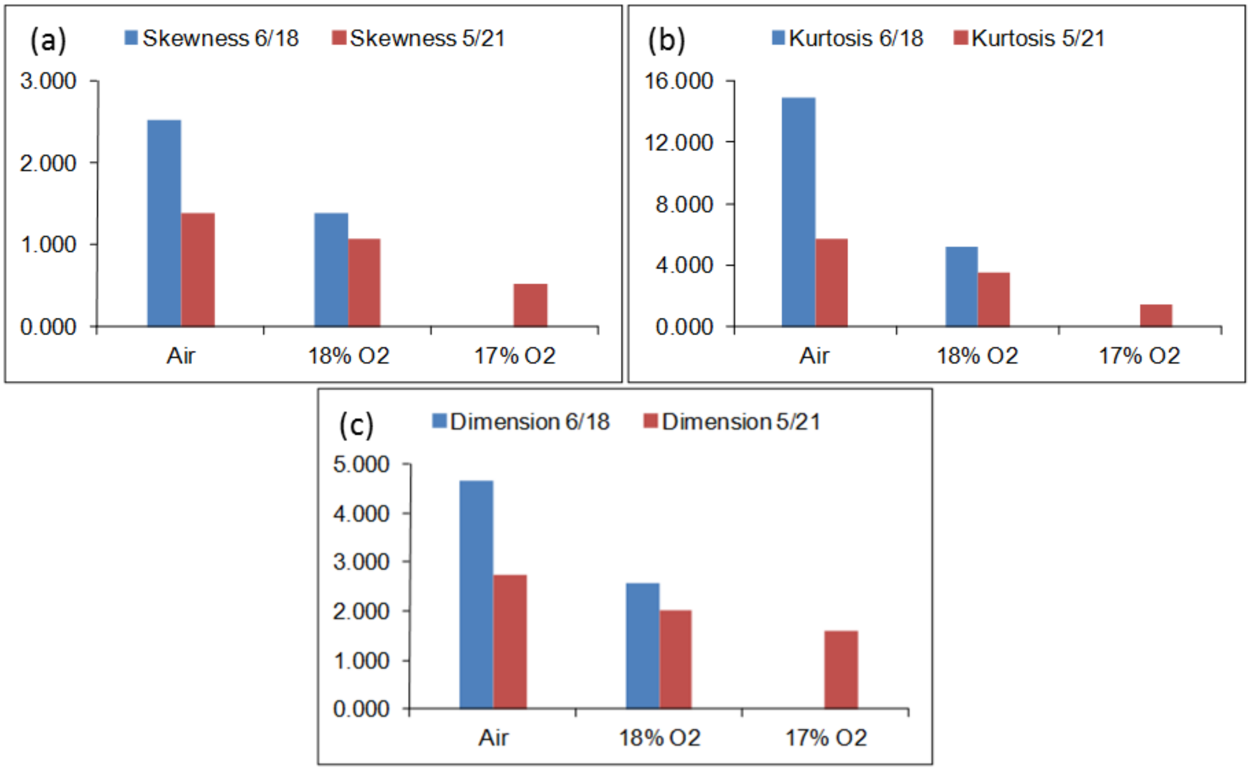


Figure 5.4. FlameDoctor measurements of (a) skewness, (b) kurtosis, and (c) dimension for flames produced by air firing and two levels of CO_2-enriched air corresponding to 18% and 17% oxygen.

5.4.2. Unburned Combustibles in Fly Ash

Combustible loss (LOI) data for the bituminous and PRB coal ashes are shown in Figure 5.5. For the bituminous ash, LOI increased with increasing CO_2 in SA (decreasing windbox O_2 concentration) ranging from 5% for air-firing conditions to 6.5% under optimum CO2-enriched conditions (18% windbox O_2 by volume). The PRB ash samples, did not show much variability between the two test conditions, ranging from 0.65% to 0.8%. The reason for this is higher char reactivity of the PRB coal relative to the bituminous coal.

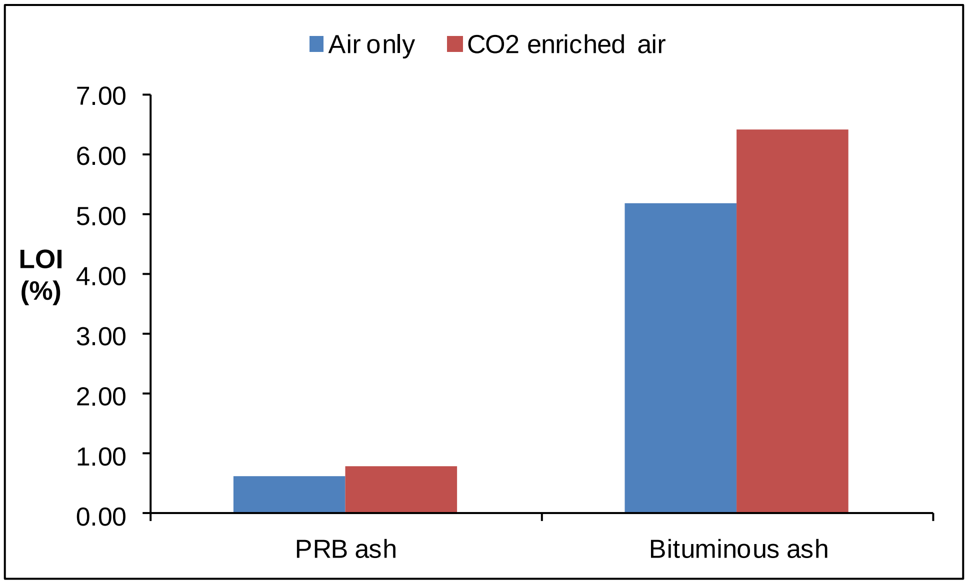


Figure 5.5. Comparison of combustible loss (LOI) for air-firing and CO_2-enriched air firing of PRB and bituminous coals.

5.4.3. Emissions

Flue gas composition was continuously measured at the convection pass exit. For tests involving bituminous coal, NO_x emissions decreased with decreasing windbox O_2, while CO levels increased (See Figure 5.6). For example, NO_x was reduced from 0.23 to 0.16 lb/MBtu between air-fired and CO_2 enriched conditions (16% windbox O_2 by volume), while CO concentration increased from 15 to 22 pm under the same conditions.

Emissions data for the PRB coal with varying windbox oxygen content (corresponding to different amounts of added CO_2) are shown in Figure 5.7. At a burner stoichiometry of 0.8, NO_x emissions were 0.14 lb/MBtu at the baseline conditions and NO_x emissions ranged from 0.13 – 0.15 lb/MBtu for CO_2 enriched tests. CO emissions at baseline conditions were about 4 ppm, and increased to 13 ppm under CO_2-enriched conditions.

Overall, this selective recycle membrane technology has a moderate effect on both NO_x emissions and CO emissions. Compared to these tests on SBS-II, one difference expected on full-scale boilers is that when CO_2-enriched air is used, both NO_x and CO emissions will be reduced as a result of re-burning mechanisms.

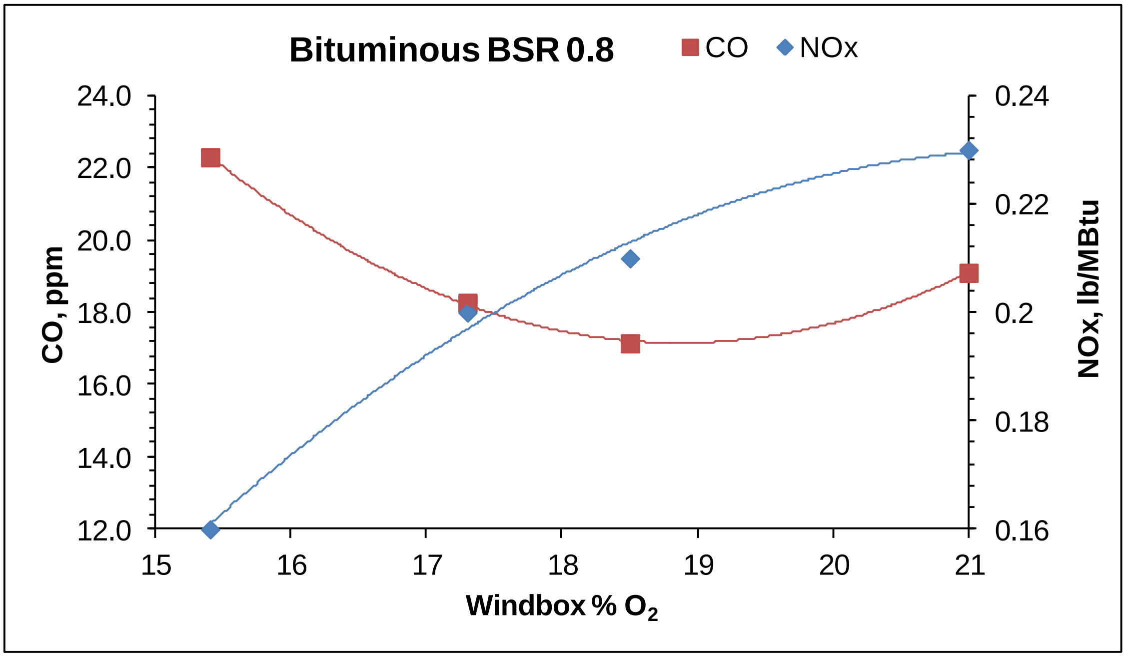


Figure 5.6. CO and NO_x emissions as a function of windbox oxygen content (corresponding to different levels of CO_2 air enrichment) for bituminous coal.

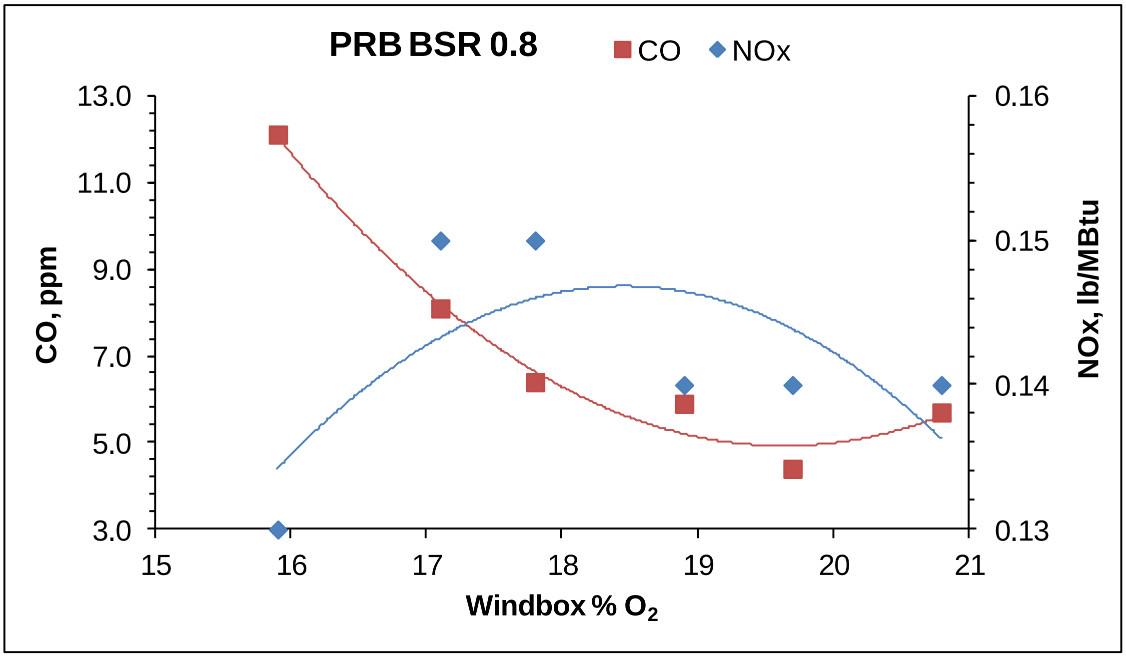


Figure 5.7. CO and NO_x emissions as a function of windbox oxygen content (corresponding to different levels of CO_2 air enrichment) for PRB coal.

5.4.4. Furnace Exit Gas Temperature (FEGT)

Furnace gas exit temperature was measured at multiple points at the top of the furnace using a high velocity temperature probe. An average value was calculated to denote FEGT under particular operating conditions. For the PRB coal tests, the measured FEGT value was about 2,270 °F under air fired conditions and about 2,030 °F under optimum CO_2-enriched (18% windbox O_2 by volume) conditions. For the bituminous coal, the measured FEGT was about 2,175 °F under air-fired conditions and about 2,020 °F under optimum CO_2-enriched conditions. FEGT data for air-fired tests is consistent with field data for high volatility bituminous coals, which tend to produce a much hotter flame than PRB coals. Boilers firing bituminous coals have a higher heat transfer in the radiant furnace, thereby lowering the FEGT. As expected, for both coals, the higher mass flow rate associated with CO_2-enriched operation produces lower FEGTs.

5.4.5. Radiant Furnace and Convection Pass Heat Absorption

One of the important goals of this project was to study the effect of CO_2 addition on heat transfer at different sections of the furnace and consequently, on boiler thermal efficiency. Data collected from heat absorption studies were used to determine boiler thermal efficiency under both baseline and optimum CO_2-enriched conditions (18% windbox O_2 by volume). There was good agreement in energy balance for both tests, as the difference between heat input and heat output was less than 3% for air-fired conditions and less than 4% for CO_2 addition tests.

Figure 5.8 directly compares heat absorption at different sections of the boiler for the two test conditions and highlights differences in heat transfer due to CO_2 addition to secondary air. Heat absorption in the furnace was higher for the air-fired test compared to the CO_2-enriched test (about 44% to 40%), and this can be attributed to higher flame temperatures under the former condition. The presence of an inert gas (CO_2) has a dilution effect and reduces flame temperature, thereby affecting heat transfer in the furnace section. Even though FEGT for the baseline condition was higher than for the CO_2-enriched condition, a higher heat absorption was achieved at the convection pass for the CO_2-enriched tests compared to the air-fired tests (35% vs 33.5% for air firing). This increase in heat absorption was mainly due to an increased mass flow of flue gas (about 20%) under CO_2-enriched conditions as compared to air-fired tests. Gas velocity increases as a result of increased mass flow and increases convective heat transfer between the flue gas and tubes.

Air heater absorption also showed a similar trend as the convection pass heat absorption, in which the CO_2-enriched test had a slightly higher transfer than the air-fired test (about 9% and 7.5%, respectively). This is mainly due to higher air heater flue gas inlet temperatures at CO_2-enriched conditions, compared to air-fired conditions. Flue gas heat loss was also higher for the CO_2-enriched test compared to the baseline air-fired test (18.5% vs 16.5 %, respectively). In commercial boilers, heat loss via flue gas is usually lower than SBS-II, as the SBS-II is designed with a partial direct secondary air heater that can provide much better flexibility for controlling flue gas temperature at WFGD/DFGD. However, heat transfer in commercial boilers is expected to follow trends similar to those observed in SBS-II.

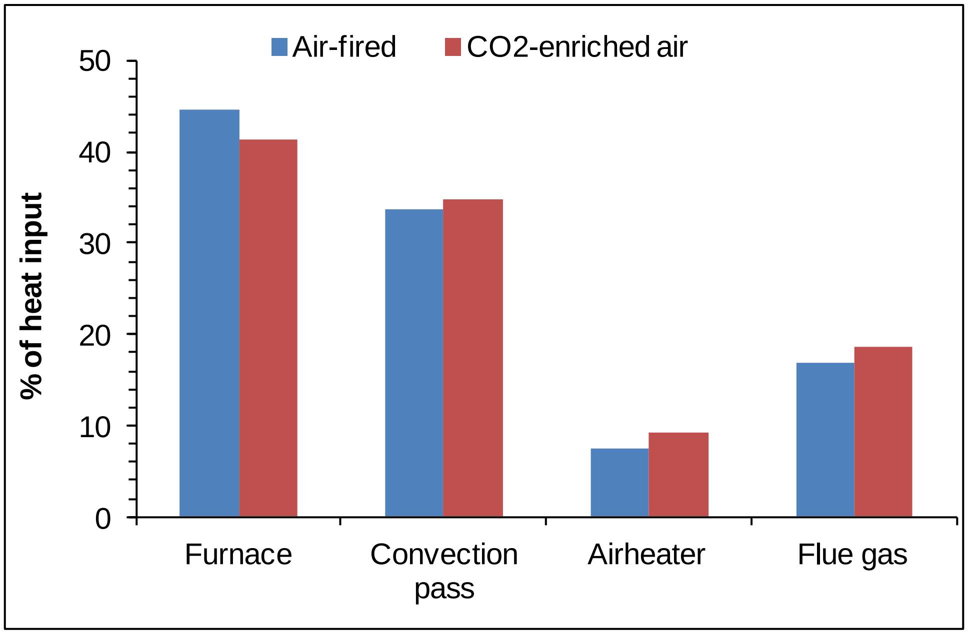


Figure 5.8. Comparison of heat absorption between air-fired and CO_2-enriched conditions at different sections of the boiler.

5.4.6. Boiler Thermal Efficiency

The overall process efficiency (thermal) was calculated using the total flue gas heat loss (sum of sensible heat and latent heat) and the heat input to the system under both test conditions. Thermal efficiency calculations will help decide if further boiler modifications are required to accommodate for CO_2-enriched conditions. Based on heat absorption data, for the air-fired conditions, η was calculated to be 83.2%, whereas for the CO_2-enriched tests at 18% windbox O_2, boiler thermal efficiency was about 81.3%. The overall difference in efficiency as a result of CO_2 addition was 1.9%. Efficiency loss calculations were also performed based on flue gas temperatures at the air heater outlet, and results indicate good agreement with heat absorption studies. Figure 5.9 shows loss in thermal efficiency for CO_2–enriched air tests compared to baseline tests as a function of windbox O_2 content. Efficiency data – both from heat absorption studies and measurement of flue gas temperature at the air heater exit – also indicate that a windbox concentration of 18% O_2 would be optimum for boiler operation without suffering a high efficiency loss.

For a supercritical boiler with a nominal net plant efficiency of 39.4%[3], an efficiency loss of 0.75% was estimated due to CO_2 enrichment of combustion air at the baseline condition (18% windbox oxygen).

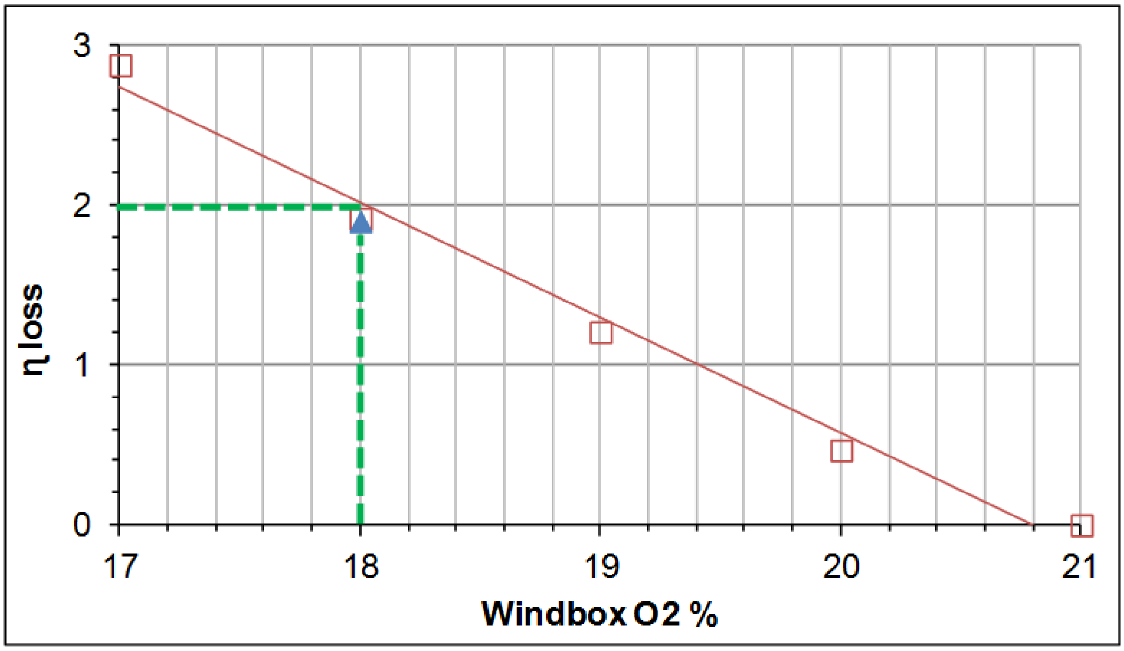


Figure 5.9. Boiler thermal efficiency loss as a result of CO_2 addition to combustion air at various windbox O_2 contents based on flue gas temperature at air heater outlet.

5.5 Conclusions and Recommendations

A CFD modeling study suggests that CO_2 recycle in secondary air is feasible as a retrofit for common boiler types firing PRB or bituminous coal provided the boiler stoichiometry is maintained by increasing the secondary air flow. Under these conditions, the predicted combustion and heat transfer characteristics for CO_2-enriched air operation showed only a modest change from the baseline air-firing results.

These modeling results were validated by testing with CO_2-enriched combustion air on B&W's SBS-II pilot boiler. The pilot boiler tests produced the following findings:

- Stable and attached flames were observed for the combustion of both bituminous and sub-bituminous coals at windbox oxygen levels varying from 21% to 16% by volume (corresponding to different levels of CO_2 recycle in secondary air). This observation was confirmed by data generated using the Flame Doctor statistical software, which showed better flame characteristics for CO_2-enriched combustion compared to conventional air firing.
- For the same burner configuration and vane settings, NO_x decreased as windbox oxygen was reduced from 21% to 16%. CO, on the other hand, showed a contrasting trend, increasing slightly from 16 ppm to 22 ppm. This general trend was observed for the combustion of the PRB and bituminous coal.
- Based on results from radiant furnace and convection pass heat absorption studies, pressure part modifications may not be required for CO_2-enriched air combustion at 18% windbox O_2.

- Boiler thermal efficiency decreased by approximately 2% at windbox of O_2 of 18%. This corresponds to an overall plant efficiency loss of about 0.75%.
- Burner windbox oxygen of 18% (corresponding to the base case MTR CO_2 recycle) was judged to be optimum for the following reasons:
    - Good flame stability,
    - reduced NO_x emissions compared to air firing,
    - minimal increase in observed unburned combustibles for the bituminous coal,
    - thermal efficiency loss is less than 2% (0.75% overall plant η loss), and
    - a relatively small change in flow rate minimizes the impact of possible tube erosion, abrasion and slagging.

Based on the positive outcome of the pilot boiler tests with CO_2-enriched air, it is recommended that:

- An evaluation of the entire membrane process for CO_2 separation integrated with a boiler should be conducted at a pilot-scale. Such an evaluation is planned as a follow-on study to this project. This new program will involve bringing the 20 TPD MTR capture unit (described in Chapter 4) from NCCC to B&W, and operating it in an integrated fashion with SBS-II.
- A full engineering study is required on a site-specific basis for future large pilots to determine individual boiler steam performance and the potential for pressure part modifications when operated with recycled CO_2.

6. TECHNO-ECONOMIC ANALYSIS

6.1. Introduction

In this section, we review the results of technical and economic studies of the MTR membrane processes. The studies performed during the program fall into two parts:

1. WP, with participation from EPRI, performed a detailed techno-economic analysis of MTR's capture process using the DOE Bituminous Baseline Study (BBS) protocol.[3] This study is included in it's entirely as Appendix A.

   In this TEA, a new full-scale supercritical coal-fired power plant (BBS Case 11) with no CO_2 capture is compared to the cost of another larger new power plant fitted with a CO_2 capture system (BBS Case 12). The second plant is sized to produce enough power to drive the capture system and to deliver the same amount of electricity to the grid (550 MW_e net). The difference in the capital and operating costs of the two plants is then taken to be the cost of electricity, which then can be compared to the amount of CO_2 captured to calculate the cost of CO_2 capture.

   WP/EPRI performed this analysis for the MTR base-case membrane process designed to remove 90% of the CO_2 produced by the second power plant, also sized to produce the same net electrical output to the grid (MTR base case). These results are then compared to BBS Case 11 (supercritical pulverized coal) and BBS Case 12 (supercritical pulverized coal with MEA CO_2 capture) with updated costs (June 2011 basis).[12]

2. In a second set of calculations, MTR performed a sensitivity study using a method consistent with the WP/EPRI approach. This analysis examined the impact of changes to the process design, system performance, and CO_2 capture rates on the capture cost.

6.2. Summary of the WorleyParsons/EPRI Study

In the WP/EPRI study, three different power plant cases were compared. All of the plants generate the same amount of electricity (550 MW_e net) to the grid. The control power plant was the BBS Case 11 supercritical power plant. This plant has no CO_2 capture system. It was compared to two other cases which incorporate different CO_2 capture systems; a mono-ethanol amine (MEA) based CO_2 capture plant (BBS Case 12), and the base case power plant equipped with MTR's membrane-based capture system developed by this program.

A variety of modeling tools were used to simulate the performance of the power plant outfitted with an MTR capture system. The membrane portion of the heat and mass balance calculation was performed using a ChemCad 6.3 process simulator modified with differential element codes for the membrane unit operations written at MTR. The overall power plant performance was simulated by WP using an Aspen Plus model initially calibrated against Case 11 of the BBS report. Boiler performance input was provided by B&W and the plant steam cycle was modeled by WP using General Electric's GateCycle software.

A block diagram of the base case membrane design is shown in Figure 6.1. The flue gas feed (stream 26) is first compressed to 2 bar (14.7 psig), after which it is sent to two membrane units in series. The first unit separates concentrated CO_2-rich permeate from the flue gas (stream 35). The CO_2-depleted residue gas is then sent to a selective recycle membrane. The bulk of the remaining CO_2 is separated by this unit into an air sweep gas circulated on the permeate side of the membrane. This air containing CO_2 (stream 10) is then used as combustion air in the power plant boiler. Because the combustion air to the boiler contains 7.6% CO_2, the boiler flue gas produced is enriched in CO_2. The result is to make the separation of a concentrated CO_2 stream from the flue gas much easier. The CO_2 concentration in the flue gas is higher than normal (close to 24% on a dry basis) and only partial removal of CO_2 by the first membrane separation unit is required, because the remaining CO_2 is removed by the selective recycle membrane.

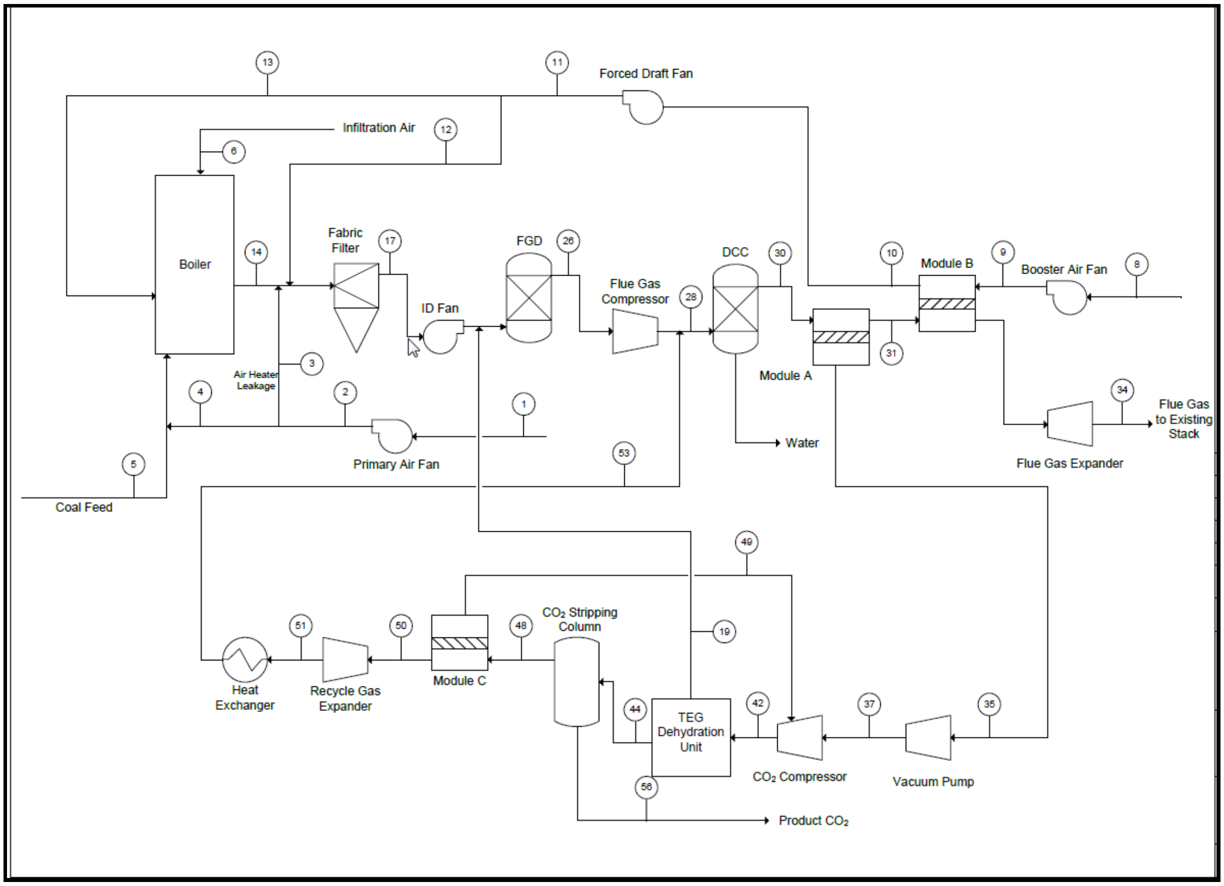


Figure 6.1. Block flow diagram of the MTR membrane CO_2 capture system.

The CO_2-rich permeate gas (stream 37) is cooled and most of the water is removed. The gas is then sent to a multi-stage compressor, which compresses the gas to 30 bar (441 psi). Cooling water is used to provide inter-stage cooling. The pressurized CO_2-rich gas is then dried with a triethylene glycol (TEG) system. The dry gas is then cooled to −30°C, with a propane refrigeration system and the bulk of the CO_2 condenses at the temperature and pressure of the gas condenser, while some oxygen and nitrogen is absorbed into the liquid CO_2. These gases are removed by sending the liquid CO_2 to a stripping column. Liquid CO_2 flows down the column, where vapor generated in the reboiler flows upward. The light gases preferentially fractionate into the vapor phase and are removed up and out of the column. High purity liquid CO_2 product (stream 56) is drawn off the bottom and pumped to the final discharge pressure of 150 bar (2,215 psi).

Overhead (stream 48) from the CO_2 stripping column still contains some CO_2. A cross-flow membrane C is used to recover this CO_2 and return it to the compression system (stream 49). The pressure differential between the column overhead and the suction pressure of the compressor stage is used as the driving force for permeation. The CO_2-lean residue gas (stream 50) that leaves the overhead membrane system is still at high pressure, so power is recovered with a low-temperature expander. The expanded residue is recycled to the flue gas prior to the direct contact cooler (stream 53). The efficiency of the propane refrigeration system is increased by using the liquid propane to reject heat to the stripper reboiler. Also, the expanded membrane residue stream, the CO_2 product, and the stripping column overheads are all at temperatures lower than that of the liquid propane. So, these gases can also be used to sub-cool the liquid propane.

The technical performance of the three cases studied is shown in Table 6.1. The BBS Case 11, no CO_2 capture plant, produces CO_2 emissions of 801 kg/MWh. The amine (MEA) plant produces emissions of 109 kg/MWh at an energy cost of 382 kW_e/tonne of CO_2 captured. 31 kW_e is extra auxiliary power used by the larger power plant, 351 kW_e is power used by the amine separation system. The membrane plant produces emissions of 111 kg/MWh at an energy cost of 358 kW_e/tonne of CO_2 captured. Of this energy, 19 kW_e is extra auxiliary power used by the larger power plant and 339 kW_e is power used by the membrane system.

Table 6.1. Summary of Technical Performance.

	BBS Case 11 (No CO_2 Capture)	BBS Case 12 (MEA)	Base Case Membrane Unit
Gross power output, MW_e	580.0	662.8	780.8
Auxiliary power requirements, MW_e (balance of plant)	30.4	47.3	40.8
Auxiliary power requirements, MW_e (capture system)	0	65.5	186.5
Auxiliary power requirements, MW_e (subtotal)	30.4	112.8	227.3
Net power output, MW_e	559.0	550.0	553.5
Net plant HHV efficiency (%)	39.3	28.4	28.6
CO_2 generated (tonne/h)	440.3	608.1	608.4
Capture efficiency (%)	0	90.2	89.9
CO_2 emitted (tonne/h)	440.3	59.7	61.2
CO_2 captured (tonne/h)	0	548.4	547.2
CO_2 emissions (kg/MWh net)	801	109	111
Energy used per tonne CO_2 captured (kWh/tonne CO_2)	0	382	358

Note: Based on 100% carbon conversion.

A comparative study of the capital and operating costs of the three cases examined is shown in Table 6.2. A more detailed breakdown of these costs is given in the WP report. Both the amine and membrane systems capture 90% of the CO_2 generated by the coal plant. The capital cost of the amine plant and the larger power plant needed is \$869 million or \$1.58 million/tonne-h of CO_2 captured. About half of this cost (or \$0.75 million/tonne-h) is the cost of the larger power plant required to drive the separation process. The remaining \$0.83 million/tonne-h is the cost of the amine separation system per tonne-h of CO_2 capture capacity. The total capital cost of the membrane system is \$783 million or \$1.43 million/tonne-h of CO_2 capture capacity, about 10% less than the amine system. Again, about \$0.75 million/tonne-h is the cost of the larger power plant, while ~\$0.68 million/tonne-h is the cost of the membrane system.

Table 6.2. Comparative Capital and Operating Costs of the Three WP Power Plants.

	BBS Case 11 (No CO_2 Capture)	BBS Case 12 (MEA)		Base Case Membrane Unit	
		Cost of Total Plant	Extra Cost of Capture Unit	Cost of Total Plant	Extra Cost of Capture Unit
Capital Costs					
Total plant cost ($ millions)	1,090	1,959	869	1,873	783
Specific plant cost ($/kW$_e$ net)	1,981	3,563	1,580	3,383	1,415
Specific plant cost ($/tonne-h operated CO_2 capacity)	0	0	1,584	0	1,431
Annual Operating Costs ($ millions/y)					
Fixed costs (labor, taxes, maintenance, etc.)	38.8	64.1	25.3 (12.9%)	61.8	23.0 (13.2%)
Variable costs (chemicals, etc.)	31.7	54.1	22.4 (11.5%)	46.3	14.6 (8.4%)
Fuel costs (coal)	104.6	144.5	39.9 (20.4%)	143.9	39.3 (22.6%)
Capital charge*	135.2	242.9	107.8 (55.2%)	232.2	92.1 (55.8%)
Total	**310.3**	**505.6**	**195.4**	**484.2**	**174.0**
Operating cost ($/tonne CO_2 captured)			42.4		37.9

* A capital charge factor of 0.124 per NETL's Cost Estimation Methodology is used for a high risk investor-owned utility.

Table 6.3 compares the cost of electricity (COE) and CO_2 capture costs for the different cases. The cost of electricity is broken down into several categories, all of which experience a slightly smaller increase in cost relative to the no capture case for the membrane process compared to MEA. The largest cost difference between the MEA and membrane processes is the variable operating and maintenance (O&M) costs. The higher MEA variable O&M cost is largely attributable to the cost of replacing solvent. Overall, at 90% capture, the COE for the membrane case is \$5/MWh$_{net}$ lower than the MEA process. Similarly, the cost of capture for the membrane process is about \$4.5/tonne CO_2 (or 8%) lower than the MEA capture case.

Table 6.3. Cost Summary from the EPRI/WP TEA.

Cost	BBS Case 11 (No CO_2 Capture)	BBS Case 12 (MEA)	MTR Membrane
Cost of electricity components ($/MWh$_{net}$)	-	-	-
Fuel	25.5	35.3	35.1
Variable O&M	7.7	13.2	11.3
Fixed O&M	9.5	15.7	15.3
Capital	38.2	73.1	70.7
Total COE ($/MWh$_{net}$)	80.9	137.3	132.3
Increase in COE compared to Case 11 (%)	-	70	64
Cost of CO_2 captured ($/tonne)	-	56.5	52.0

In comparing these TEA results with prior studies, there is general agreement that the MTR membrane process is competitive with solvent capture systems at 90% capture. MTR and WP used a conservative cost estimating philosophy in preparing this TEA. With the exception of the membrane modules, the balance of process equipment selected by WP (representing a significant fraction of the capital and operating costs), are commercial equipment that are available today in the required sizes. In that regard, the calculated COE and capture cost values presented here more closely relate to a first of a kind plant than an nth of a kind plant. For comparison, a prior DOE study (the "Pathways" report)[4] examined MTR's capture system utilizing advanced compression/vacuum equipment and low risk technology financing and found capture costs of less than $40/tonne CO_2.

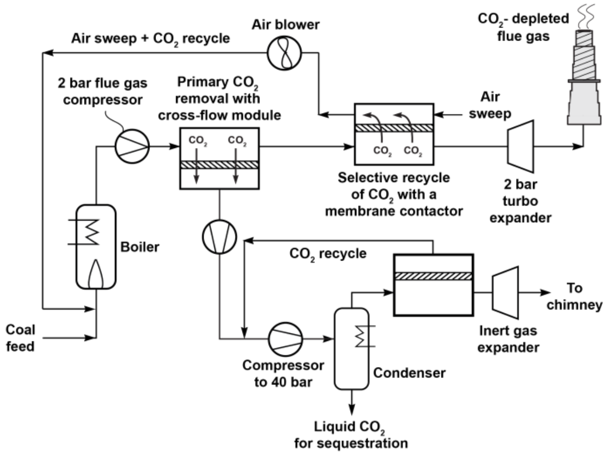
Another difference between the Pathways study and the current TEA is the feed pressure used in the membrane design. The results shown in Tables 6.1, 6.2, and 6.3 are for a membrane system where the flue gas feed is compressed to 2.0 bar. This elevated feed pressure helps to reduce the membrane area and the size of the capture plant. However, this is accomplished by large, costly compressors using significant energy. For example, about 25% of the energy use for the base case membrane system described in Table 6.1 is for feed compression. In contrast, the DOE Pathways report looked at minimal feed compression (blower only), and found the potential for lower costs if high performance membranes are used. Better understanding the impact of membrane system design variables (such as feed pressure) on capture costs is a useful objective.

In addition to feed pressure, another variable that is expected to play a significant role in impacting the cost of capture is the capture rate. Membrane separation systems are well-known to be particularly cost-effective for bulk removal applications. While most of the CO_2 capture literature has focused on 90% capture, membranes are likely to be very attractive for lesser removal rates. Understanding the impact of capture rate on cost seems especially relevant given proposed CO_2 emissions regulations (such as the EPA Clean Power Plan) that amount to partial capture from coal-fired plants (for example, a proposed emissions limit of 1,400 lb CO_2/MWh amounts to <30% capture from an average coal-fired power plant).

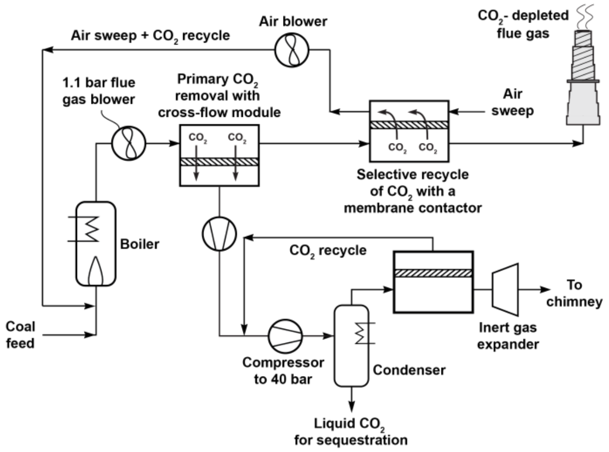
6.3. Sensitivity Study Cases

Based on the issues described above, MTR conducted a sensitivity study to examine the effect of membrane process design and CO_2 capture rate on the cost of capture. Figure 6.2 shows the base case membrane design [Figure 6.2(a)], as well as two possible variants. The design shown in Figure 6.2(b) has the same form as the base case, but the flue gas feed pressure is reduced to 1.1 bar (1.5 psig). The large flue gas compressor of the base case is then replaced with a much less energy-intensive blower and a flue gas turbo expander is no longer used. The third design, shown in Figure 6.2(c), does not include selective recycle to the boiler generated by the membrane contactor. Elimination of this recycle membrane simplifies the process flow scheme, but reduces the CO_2 concentrations in the flue gas stream to be treated. This makes it more difficult to efficiently achieve higher CO_2 capture rates from the flue gas. For this reason, the no recycle case is limited to CO_2 capture in the 20 to 50% range.

(a) Base Case - 2 bar feed with selective recycle to boiler (90% CO_2 capture)



(b) Low Energy Case - 1.1 bar feed with selective recycle to boiler (40-80% CO_2 capture)



(c) Low CO_2 Capture Case - 1.1 bar feed, no selective recycle (20-40% CO_2 capture)

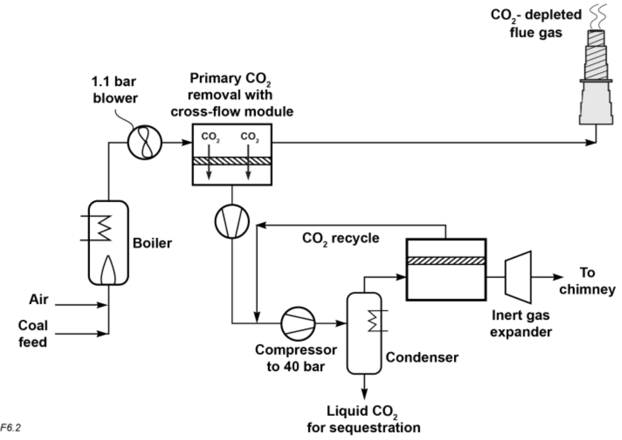


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Figure 6.2. Three potential membrane process designs: (a) The Base Case design shown in Figure 6.1 (feed pressure 2 bar, permeate pressure 0.2 bar). (b) The Low Energy Case (feed pressure 1.1 bar, permeate pressure 0.2 bar). (c) The No Recycle Case (feed pressure 1.1 bar, permeate pressure 0.2 bar, no selective recycle to the boiler).

The calculated energy consumption and membrane area used by the three processes are very much a function of the CO_2 removal from the flue gas (see Figure 6.3). For all three designs, the membrane area used per tonne-h of CO_2 removal through the membrane increases as the CO_2 capture rate increases. This is because the CO_2 concentration in the feed flue gas decreases as more CO_2 is removed through the membrane. As a result, more membrane area is needed for each incremental increase in CO_2 removal because the driving force for CO_2 removal is being reduced. The CO_2 concentration in the permeate stream also decreases, and the vacuum and compression pumps required to bring the CO_2 and an increasing amount of non-condensables to 30 bar where they can be removed all become larger.

All three designs show a minimum in the energy consumption required per tonne of CO_2 capture. This reflects the trade-off between feed compression energy per tonne of CO_2 that decreases as the CO_2 capture rate increases; and the energy consumption of the permeate vacuum and compression steps that increase as the CO_2 removal rate increases and the concentration of CO_2 in the permeate gas falls.

As shown in Figure 6.3, in the 2 bar base case design, the trade-off described above results in a minimum energy at close to the DOE target value of 90% CO_2. At 90% CO_2 capture, the energy consumption of the 2 bar design is higher than the 1.1 bar design, but this disadvantage is somewhat offset by the 2 bar design's much lower membrane area. High permeance and/or low cost membranes will minimize this lower membrane area advantage. Figure 6.3 also shows that a significant reduction in the energy cost of CO_2 capture is possible if the CO_2 removal rate is lowered to the 40 to 70% range. In this range, the energy consumption of the 1.1 bar design is 60 kW_e/tonne CO_2 lower than the 2 bar process minimum design. The 2 bar case still uses less membrane area, but this advantage is much reduced. These results indicate that the 1.1 bar feed case is preferred for most capture rates, particularly at less than 90% capture.

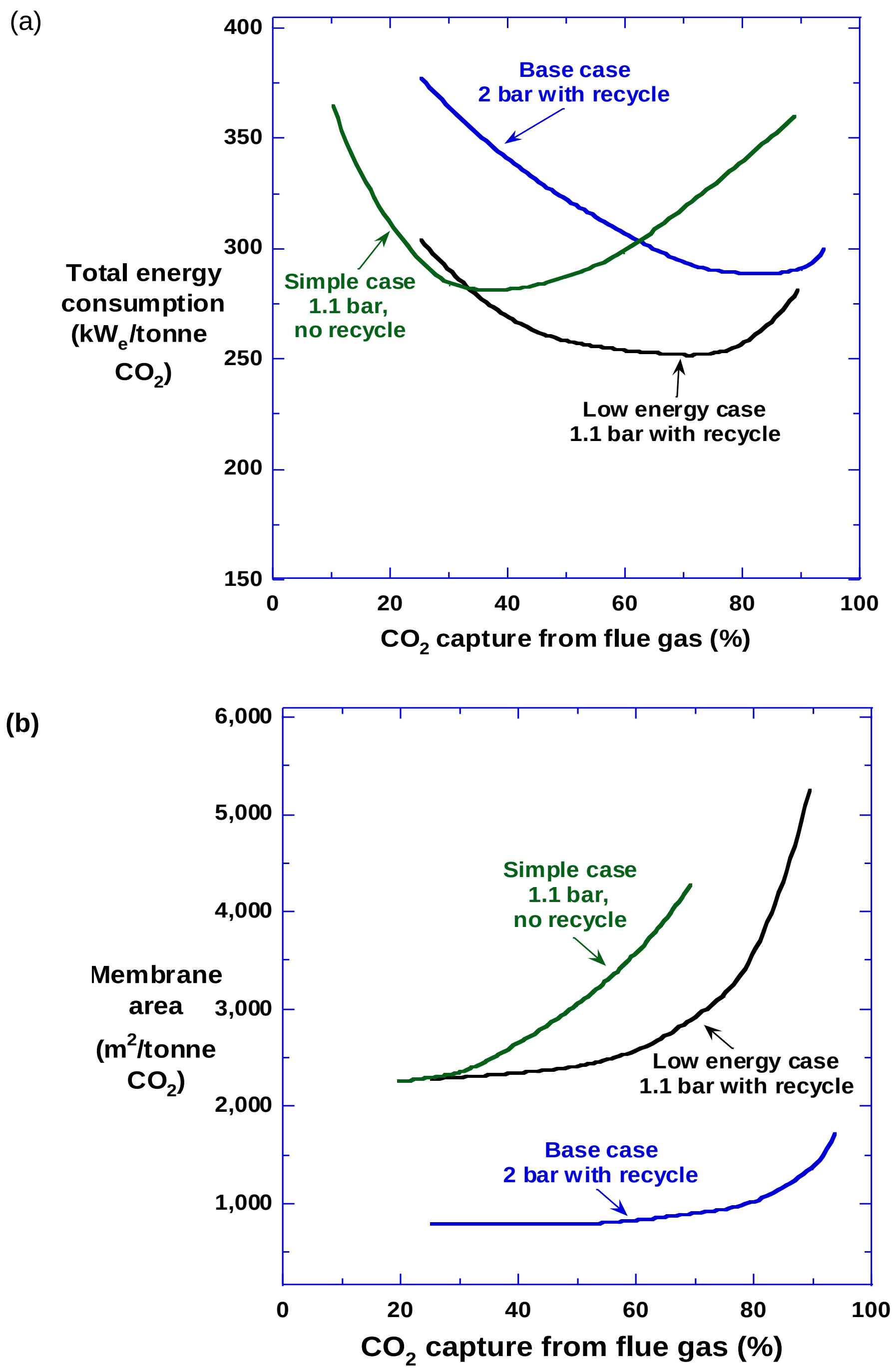

Figure 6.3. (a) Total energy consumption and (b) membrane area required per tonne of CO_2 captured as a function of fractional CO_2 capture from flue gas.

The last design shown in Figure 6.3 is for a low feed pressure system (1.1 bar) with no selective recycle membrane; and hence, no recycle of CO_2 to the boiler. Energy is then only used to circulate flue gas through the first-stage membrane unit, so the energy consumption of the initial feed separation step is low. Because the contactor is not used to recycle CO_2, there is no enrichment of CO_2 in the flue gas. This means the CO_2 concentration in the membrane permeate is relatively low, especially at high CO_2 capture rates. As a result, the energy consumed by the permeate vacuum and permeate compression system is high. For this reason, the energy use of the no recycle case increases dramatically for capture rates of greater than 50%.

A major advantage of the no recycle design is that no changes to the power plant boiler are required. Power plant boilers are massive operations and are carefully optimized for maximum efficiency. Although our studies with B&W indicate that recycle to the boiler is viable, it seems likely that there will be many cases where plant operators are reluctant to interfere with the operation of these units. That the simple no recycle case shows a minimum energy in the 30% capture range is advantageous because this capture rate corresponds with the EPA's proposed initial limits on plant CO_2 emissions.

6.4. Cost Comparison

Energy consumption is an important factor determining the feasibility of a flue gas separation process, but the ultimate figure of merit is the capture cost (\$/tonne CO_2 captured). The capture cost is a combination of labor, maintenance, membrane replacement and electric power, but also contains a charge to cover the capital cost of the plant. Based on the performance data shown in Figure 6.3, the capture cost as a function of capture rate was calculated for the two of the membrane process designs: the low energy 1.1 bar feed case, and the simple no recycle case. The key assumptions used in these calculations are shown in Table 6.4. The compressor, vacuum pump, and turbo expander efficiencies and costs are average values for large gas processing systems. The membrane skid cost of \$50/m^2 and module replacement cost at \$15/m^2 reflects the costs already achieved in large reverse osmosis plants where the area of membrane needed is similar to that required for a membrane system capturing 90% CO_2 from a 500 MW$_e$ coal-fired power plant.

Table 6.4. Assumptions Used in Design Calculations.

Category	Value ($)	Units
Polytrophic compressor, turbo expander, and vacuum pump efficiency	85%	-
Compressor and vacuum unit cost	700	\$/kW$_e$
Refrigeration/cooling system cost	1,000	\$/kW$_e$
Turbo expander cost	1,000	\$/kW$_e$
Membrane CO_2 permeance	2,500	gpu
Membrane CO_2/N_2 selectivity	50	-
Replacement membrane module cost	15	\$/m^2
Membrane skid cost (including initial fill of modules)	50	\$/m^2
Equipment installation factor	100%	-
Capital depreciation/interest	15	%/year
Cost of power	0.05	$kW

In 2015, MTR commissioned the Trimeric Corporation to perform a study of the cost of vacuum and compression equipment for a full-scale membrane flue gas CO_2 capture system. The study was particularly focused on the cost of the feed compressor unit required to compress the flue gas from 1.0 to 1.5 bar, and the vacuum equipment required to take the permeate CO_2 rich gas from

0.2 bar to 1.0 bar. This type of equipment is not widely used with flue gas or at the scale needed for this process. Because equipment required to compress the CO_2 stream from ~2 to 150 bar is available and has been studied, MTR did not include this equipment in Trimeric's study.

For the flue gas compression equipment, vendor quotes ranged from \$340-1,400/ kW_e. The lowest cost was for a very large 52 MW_e axial fan machine. However, the best option seemed to be multi-stage integrally geared machines in the 8 to 10 MW_e range, costing \$500-800/kW_e (including motor drives). These machines had polytropic efficiencies of 84 to 89%.

Vacuum equipment (0.2 to 1.0 bar) was priced in the \$650-1,400/MW_e range. Axial compressors seemed the best option, with claimed polytropic efficiencies of 88 to 90% and costing ~\$650-700/kW_e. Individual machines would be in the 7-10 MW_e range, so 3 to 6 machines would be needed for a 550 MW power plant.

The installation cost factor in Table 6.4 is 100% of the sum of all the large equipment items (Bare Erected Cost). This factor is less than what is normally used in petrochemical/refinery plants and needs some explanation. Most petrochemical/refinery unit operations follow the 0.6 power rule to achieve economies of scale. The cost of a unit operation increases in proportion to its production rate raised to power 0.6. Thus, a tenfold increase in production rate increases cost by $10^{0.6}$, or approximately four-fold. Membrane technology has not followed this path. Membranes are packaged as individual modules (elements) typically containing 20 to 100 m^2 of membrane. A membrane plant has many of these modules manifolded together. For example, a membrane plant removing CO_2 from natural gas may contain as many as 10,000 individual modules, arranged in tubes, 6 to 7 modules/tube, 30 to 40 tubes/skid, and 40 to 50 skids for the whole plant. Economics of scale are achieved through volume production methods. Membrane is made in 5,000 meter rolls, and automated robotic equipment is used to make individual modules. Many identical membrane skids are then fabricated in a large machine shop. The skids are sent to the site as prefabricated container-sized units designed to be linked together with minimal on-site work (see for example, Figures 4.5 – 4.8 showing assembly of the 20 TPD small pilot system). A significant benefit of this approach is that almost all plant fabrication is performed in a well-equipped and well-controlled factory environment. On-site field fabrication is hard to control and is typically two to three times more expensive than the same work performed in a fabrication shop.

Large membrane plants are also built as a series of separate trains arranged in parallel, each train equipped to operate independently of the others. A 500 MW_e coal power plant might be equipped with up to ten identical trains, each able to treat the flue gas generated when producing 50 MW_e of power. The total compression energy required by a single 50 MW_e train is about 10 to 12 MW_e. These are large machines, but small enough to be transported to the site on one or two large trucks.

By using multiple identical modular trains; design, fabrication and installation costs are significantly reduced. Multiple trains also make turn-down and turn-up easier, reduce the inventory of spare parts needed and eliminate the need for redundant components required for large single-train plants. For the reasons cited above, an installation factor of 100% of the cost of the major equipment items is sufficient to cover site work, assembly and connecting piping and electrical systems for a membrane CO_2 capture plant.

Based on these assumptions, the cost of CO_2 capture as a function of fractional CO_2 recovery is shown in Figure 6.4 for two cases:

1. The two-step selective recycle process (blue curve), and
2. A single step membrane process without recycle to the boiler (black curve).

The data in Figure 6.4 demonstrate two important points. First, the cost of capture for a membrane system is not constant with capture rate, and in fact, shows a minimum at <70% capture, the exact value of which depends on the process design. For the base-case 2-step process with selective recycle, the minimum capture cost of ~$35/tonne CO_2 occurs around 60% capture. At 90% capture, the cost for this process is on a steeply increasing trend. The second point is that for a single step membrane process without selective recycle, the capture costs are much higher at 90% capture (>$70/tonne). In fact, the difference between the cost curves for the two processes shows the benefit of selective recycle, which averages about $20/tonne at capture rates >70%. However, at lower capture rates (for example, between 20 – 40%), the difference between the capture costs for the two processes is very small. Considering the simplicity of the one step process without recycle, this may be the preferred approach for a membrane capture system that meets EPA proposed CO_2 emissions limits.

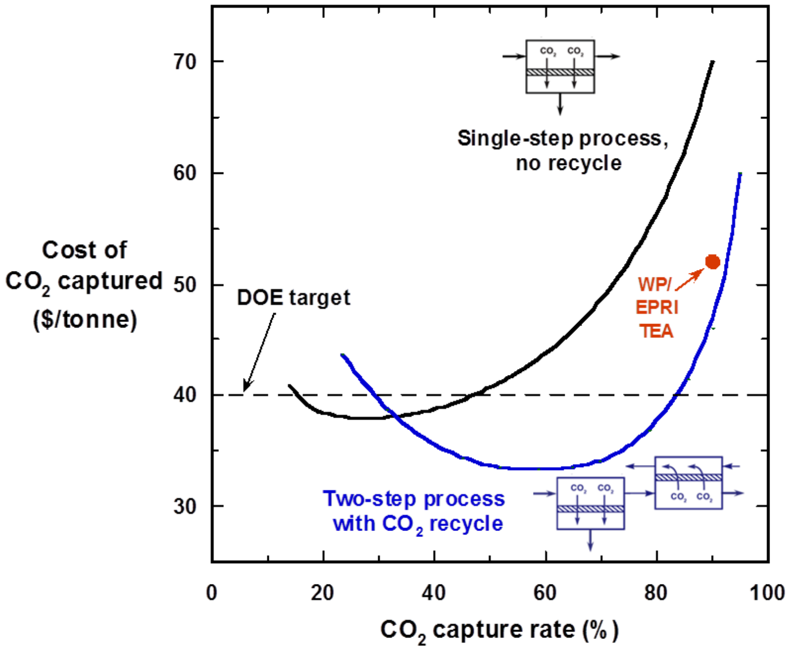

Figure 6.4. Cost of capture as a function of capture rate for two variations of the MTR membrane process. The blue curve is calculated for the baseline two-step design with selective CO_2 recycle. The black curve is for a one-step, no recycle design. Also shown in the plot is the data point calculated by WP/EPRI at 90% capture.

7. FIELD DEMONSTRATION OF ADVANCED MEMBRANES FOR SYNGAS CLEANUP AND CO_2 CAPTURE

7.1. Introduction

This chapter describes project work to evaluate membranes for industrial CO_2 capture. While the power industry produces the majority of non-transportation man-made CO_2 emissions, there are other large industrial CO_2 emission sources. For example, steel production, cement manufacture, and chemical refineries each generate between 5 and 10% of worldwide CO_2 emissions from stationary sources. Unlike the power industry, there is no renewable or decarbonized means of producing these valuable industrial materials. As a result, CO_2 capture options will be needed to reduce emissions from industrial sectors. Membranes are an attractive industrial capture approach for several reasons. For example, industrial CO_2-containing streams often exhibit a higher CO_2 partial pressure than power plant flue gas. This makes CO_2 capture easier, particularly for membranes which separate species based on partial pressure differences. In addition, membranes do not use steam (in contrast to solvent capture processes), which may not be as readily available in industrial settings as it is in a power facility.

During this project an industrial capture test opportunity was identified at a municipal waste-to-biofuels facility operated by Enerkem near Edmonton, Canada. Design calculations showed that two types of MTR membranes could be used for CO_2 separation and process efficiency improvements:

- a CO_2-selective Polaris membrane optimized for high pressure, cold operation would perform CO_2/H_2 separation on syngas leaving the bio-waste gasifier to debottleneck or replace a Rectisol separation system;
- a H_2-selective Proteus membrane can recover H_2 from a H_2/CO_2 mixture, leaving a methanol reactor to improve the efficiency of the biofuels production process.

Both of these membrane types have previously been developed with DOE support. The history of the Polaris membrane has been discussed in Chapters 1 and 2. For this industrial field test, the Polaris Gen-1 membrane was modified to maximize CO_2/H_2 selectivity. This effort involved (1) reducing membrane CO_2 permeance because permeance is not as important for high pressure syngas treatment as it is for flue gas CO_2 capture, and (2) adapting the membrane structure for sub-ambient temperature operation (~0°C) where selectivity over hydrogen is enhanced. The Proteus membrane was originally developed for pre-combustion CO_2 capture under DOE award number DE-FE0001124. This membrane has uniquely high H_2/CO_2 selectivity and has previously been field tested with coal-derived syngas at the NCCC. The capture program described in this chapter will be the first full-length module test for these membrane types at an industrial site.

For this test, a bench-scale membrane system (approximately 1 TPD) was designed and built by MTR. Preparation work at the industrial host site (the Advanced Energy Research Facility or AERF) was co-funded by Alberta Innovates-Energy and Environmental Solutions (AI-EES) and performed by MTR's collaborators, Enerkem, and the City of Edmonton (CoE). In addition to site preparation, EK and the CoE will aid in the design and installation of the membrane skid, construct process tie-in lines, and manage syngas plant operations.

7.2 Background

The objective of this program is to demonstrate successful, industrial pilot-scale use of newly-developed advanced membranes for the separation of CO_2 from industrial processes. The increasing concern about the impact of anthropogenic emissions of CO_2 on global climate is a key driver for this work. A number of states and provinces, including Alberta, Canada, have implemented, or are considering implementing a carbon tax, cap-and-trade, or other incentive program to reduce the carbon footprint of power production and industrial processes that emit large quantities of CO_2. These regulatory and/or legislative actions are driving industry to explore strategies to reduce CO_2 emissions, including efficiency improvements and direct capture of CO_2 from emission sources.

A variety of separation technologies are being considered for CO_2 capture from large point sources, including absorption, adsorption and membranes. Gas separation membrane technology has emerged as a standard industrial unit operation over the past 30 years. Today, gas separation membranes are widely used for natural gas sweetening (acid gas removal) and hydrogen recovery in refineries. Among the advantages of membrane systems are environmentally-friendly operation (no hazardous chemical handling, disposal or emission issues), relatively low energy use, simple design and operation, low maintenance costs and compact size. Some of the drawbacks in using a membrane separator are difficulty in achieving high purity and recovery, and a lack of operating experience in many proposed industrial process applications, which fosters concerns about reliability. Overcoming these drawbacks through development of advanced membranes with better separation performance, and field testing such membranes to demonstrate reliability is an ongoing effort in the membrane industry.

Recently, MTR – with funding from the DOE – has developed advanced membranes for CO_2 capture. Two membrane types have been brought to the early field demonstration stage:

- *Polaris membranes.* These membranes have the unique ability to selectively permeate CO_2 over H_2 (and other light gases). Conventional membranes are good at removing CO_2 from CH_4, N_2, or CO, but are typically slightly H_2-selective over CO_2. Polaris membranes allow CO_2 to be stripped from syngas streams, while leaving the valuable H_2 in a high-pressure residue. They can be used in combination with or to replace conventional acid gas removal technologies such as Selexol™ or Rectisol®.
- *Proteus membranes.* These new membranes have H_2/CO_2 selectivity greater than 20, compared to conventional commercial membranes with H_2/CO_2 selectivities of less than 5. At a slightly earlier stage of development than Polaris, Proteus membranes show potential for recovering relatively pure H_2 from syngas streams, including from reformer operations and IGCC power production.

While MTR membranes are showing progress toward long-term use for CO_2 capture from power facilities, there are also near-term opportunities to demonstrate relatively low-cost CO_2 capture from industrial sources with membranes. In some cases, the captured CO_2 may have economic value for use in Enhanced Oil Recovery (EOR) operations or as feedstock for chemicals production. The critical next step will be to identify these industrial opportunities, demonstrate membrane operation with commercial-sized components, and perform an economic evaluation of the process. This is the aim of the current test program.

7.3 Host Site Information

Finding a host site to conduct an industrial CO_2 capture test proved more difficult than originally anticipated. During budget period 1, discussions were initiated with several candidate host sites in the U.S. However, with limited funding for site preparation and test utilities, and no imminent industrial CO_2 emission regulations, we were unsuccessful in getting a test site commitment. Finally, in early 2013 an opportunity surfaced to conduct a test within a waste-to-biofuels facility near Edmonton, Canada.

Figure 7.1 shows an aerial view of the Edmonton Waste Management Centre. This 550 acre facility processes municipal waste from the City of Edmonton, recycling or converting to products 90% of the waste input. Circled in the overview picture are the Enerkem Alberta Biofuels (EAB) plant and the Advanced Energy Research Facility (AERF). The EAB plant is a commercial facility that produces up to 10 million gallons/year of methanol (or ethanol) from biomass using Enerkem’s gasification and conversion technology. The AERF is a pilot test facility co-owned by the CoE and AI-EES, and operated by Enerkem. It is dedicated to the development and demonstration of innovative technologies to convert waste to clean energy and products. Figure 7.2 shows a picture of the test bay inside the AERF.

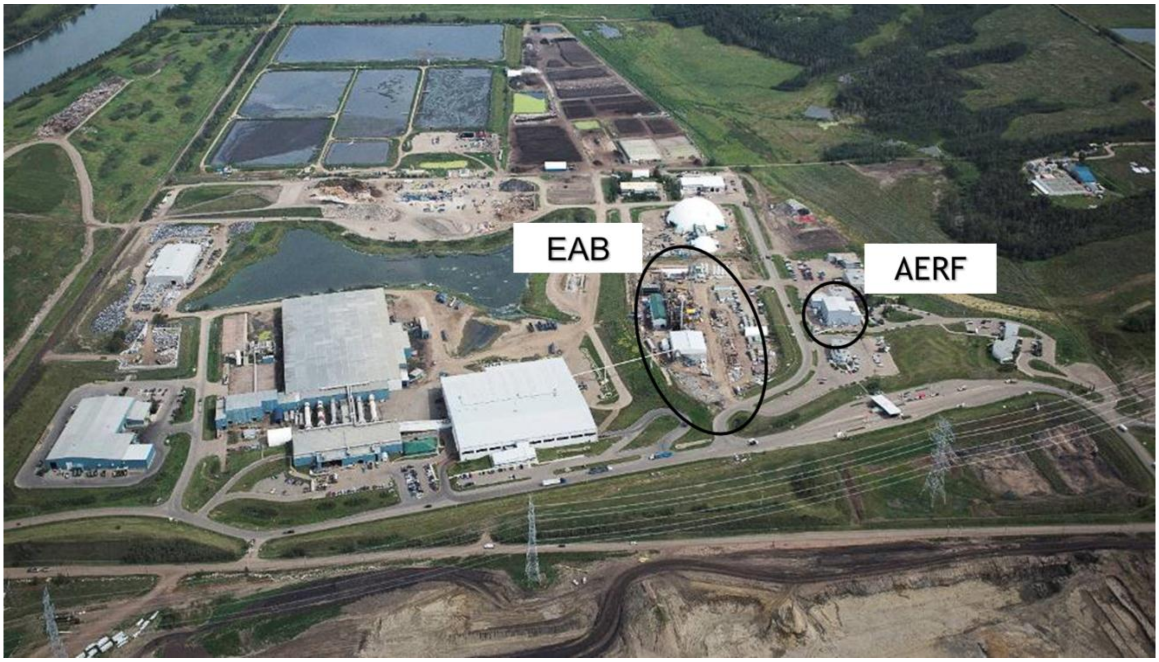

Figure 7.1. An aerial picture of the Edmonton Waste Management Centre. The locations of the Enerkem Alberta Biofuels (EAB) plant and the Advanced Energy Research Facility (AERF) are circled.

Figure 7.2. A picture of the test bay at the AERF. This 300 kg/h pilot facility is dedicated to demonstrating innovative technologies for converting waste to clean energy and products.

Based on discussions with Enerkem personnel, it was determined that advanced MTR membranes could be beneficial in improving the efficiency and reducing CO_2 emissions from the Enerkem waste-to-biofuels process. CoE and AI-EES agreed to provide funding for site preparations at AERF, and in October 2013 it was agreed that this site would host a 1,000 hour test of MTR membranes treating industrial syngas produced by the EAB plant.

7.4 Summary of Industrial Test Program Accomplishments

Initially, we hoped to build the industrial test skid, install it at AERF, and finish the 1,000 hour test by summer 2015. In the end, this proved to be too ambitious due to a number of delays at the host site mostly related to supply of syngas from the EAB. At the date of this report, the completed test skid is installed at the AERF, but testing has been pushed back to the summer of 2016. We now plan to issue an update to this report once the test data analysis is completed later this year. A summary of the completed industrial capture test work is described below organized by subtask.

7.4.1 Design Membrane Test Skid

The membrane skid designed by MTR engineers will be treating two industrial gas streams during operation. A block flow diagram showing where the MTR skid will be integrated within the EK pilot process is shown in Figure 7.3. A slipstream of conditioned syngas (sulfur and particulates removed), will be provided by the EAB plant located next to the AERF via an underground pipe. A portion of the EAB syngas feed will be directed to the Polaris membranes (stream 1). The Polaris residue and permeate streams (2 and 3) will be combined and recycled upstream of the membrane skid. The off-gas from the 3-phase methanol reactor will be sent to the high-temperature Proteus section of the skid (stream 4). The Proteus residue stream will be sent to the thermal oxidizer (TOx) or directed to a reformer (ATR) for further processing. The Proteus permeate (stream 6) can optionally be sent to the TOx. The mass and energy balances showing the anticipated performance during nominal test conditions are given in Table 7.1. The following summarizes the skid design work completed in preparation for construction:

- Mass and energy balances based on design conditions provided by EK were finalized.
- Vessels and skid equipment were sized and determined to be CRN-compliant (Canadian Registration Number, similar to AMSE in the U.S.).
- Instrumentation was selected and ordered.
- Vessel drawings were submitted to the Alberta Boiler Safety Association (ABSA) for approval and CRNs were obtained for both the membrane and heater vessels.
- All electrical drawings were prepared and completed for the Programmable Logic Controller (PLC) and heater control panels, as well as the heat tracing lines in accordance with the Canadian Electrical Standard (CSA).
- Final, pre-fabrication P&ID's were reviewed internally and with project partners.
- A Hazard and Operability Study (HazOp) review of the membrane skid was completed with MTR, EK and CoE personnel.
- Final, pre-fabrication general arrangement (GA) and isometric drawings for the membrane skid were prepared and submitted to the skid fabricator (Johansing Iron Works, Oakland, CA).

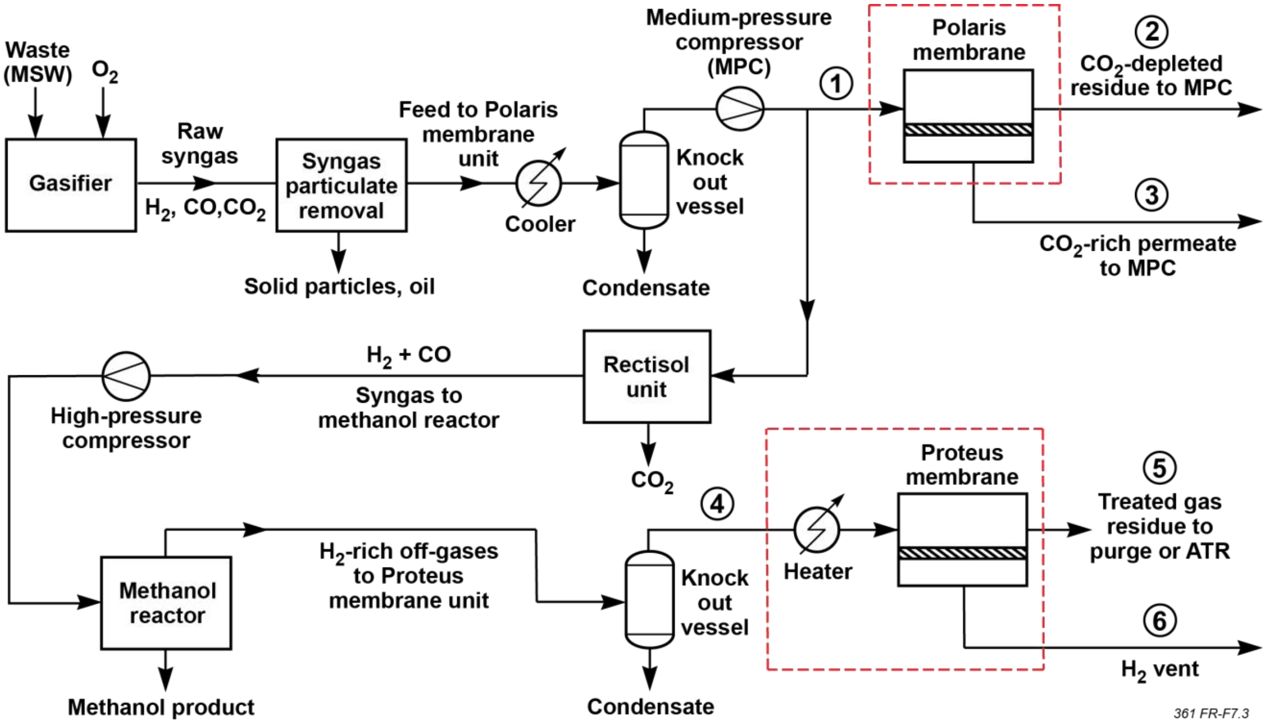

Figure 7.3. Industrial CO_2 capture process flow diagram. Streams designated 1-6 are tie-in points between the EK process and the MTR membrane skid. The dashed red boxes represent the process boundaries of the MTR membrane units (both membrane units mounted on the same skid).

Table 7.1. Mass and Energy Balances Reflecting the Anticipated Performance of Polaris and Proteus Modules during Operation.

Stream No.	①	②	③	④	⑤	⑥
Molar flow (kmol/h)	7.41	5.93	1.48	7.55	5.28	2.27
Mass flow (kg/h)	149.10	100.6	48.45	105.90	98.47	7.46
Temperature (°C)	15.00	11.64	13.32	35.00	151.40	150.70
Pressure (bar)	25.81	25.71	1.20	63.11	63.01	1.20
Components (mol%)						
Hydrogen	42.43	47.63	21.59	51.17	32.26	95.24
Carbon monoxide	20.40	23.86	6.58	21.27	29.61	1.82
Carbon dioxide	25.79	15.48	67.02	6.77	9.18	1.14
Methane	8.15	9.37	3.25	14.92	20.77	1.28
Nitrogen	2.08	2.49	0.43	3.73	5.19	0.32
Water	0.08	0.02	0.32	0.00	0.00	0.00
Methanol	0.00	0.00	0.00	0.62	0.84	0.10
Ethylene	1.07	1.14	0.81	1.53	2.14	0.09

7.4.2 Skid Construction and Membrane Module Preparation

A summary of the skid construction activities is as follows:

- A factory acceptance test (FAT) of the programmable logic controller (PLC) panel was performed by MTR personnel at the PLC fabricator (Industrial Equipment Solutions, Corona, CA).
- The membrane skid frame was constructed and painted.
- The three vessels (two membrane module vessels and the heater vessel) were fabricated and honed by MTR's local fabricator.
- ABSA approvals of the module vessels and the heater vessel were acquired. The CRN stamps were obtained and affixed to the fabricated vessels.
- All skid instrumentation and piping was procured. The PLC panel and heater panel were delivered to the skid fabricator.
- Full assembly of the skid (vessels, piping, instrumentation, and electrical control panels) was completed at the fabricator.
- Electrical tie-in of all internal instrumentation to the PLC and heater cabinets was completed by a local contractor (Zeco Electric).
- Commissioning of the PLC programming and internal electrical tie-ins (loop checks) was performed at Johansing Iron Works by MTR personnel.
- The PLC control documents (interlock and sequence descriptions) and skid electrical drawings were finalized by MTR personnel.
- Hydro-testing of the membrane skid was performed by the fabricator and witnessed by MTR personnel.
- Vessel heat tracing and insulation were installed at the fabricator.
- The as-built skid P&ID and general arrangement (GA) drawings were finalized and distributed to the project partners.

The layout of vessels and instrumentation on the membrane skid is shown in the as-built GA drawing (Figure 7.4). Flanged tie-in points for EK process connections are located at the two ends of the skid (upper right and lower left of the GA drawing). The horizontal membrane module vessels that will house the Polaris and Proteus modules are labeled in the figure.

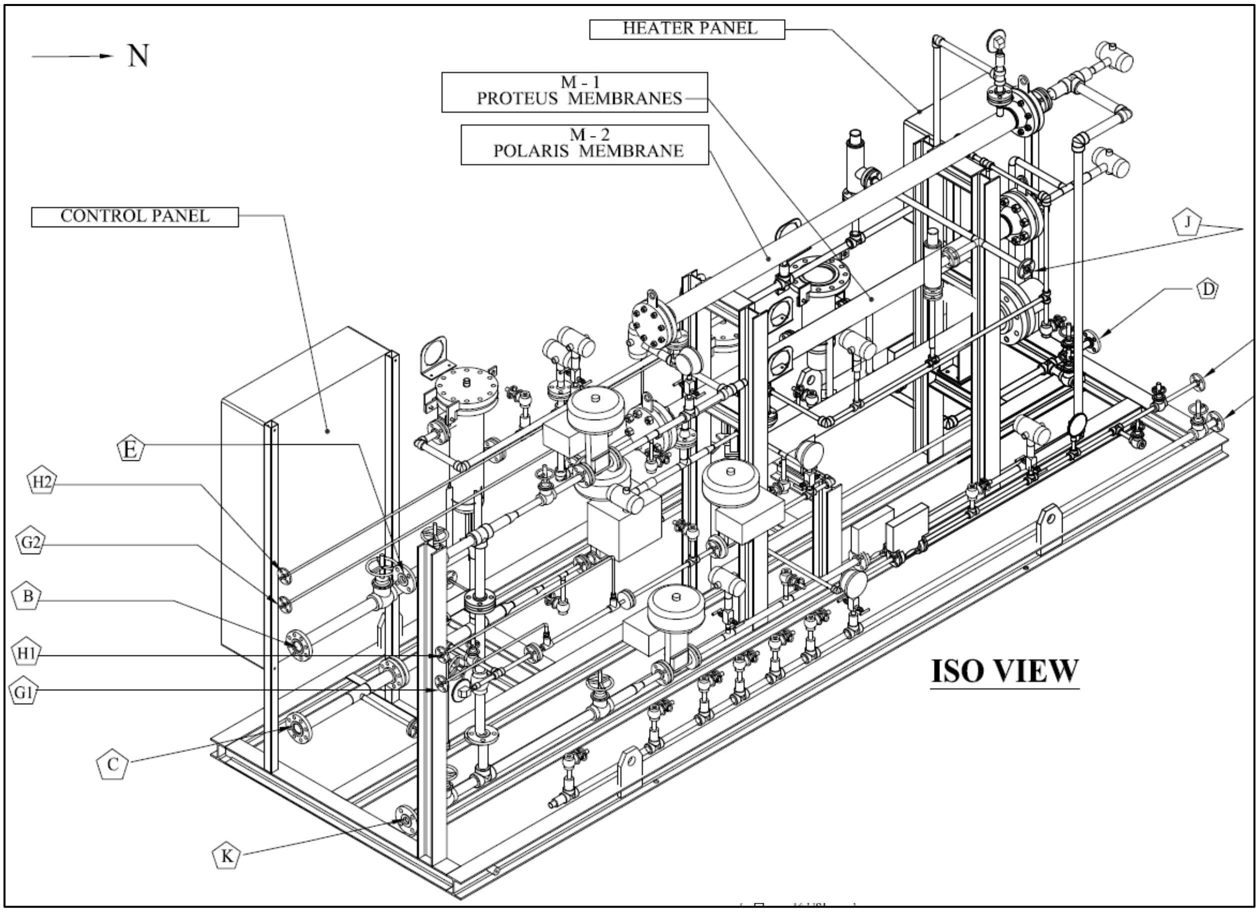

Figure 7.4. As-built general arrangement drawing of the MTR membrane skid.

Concurrent with the skid construction activities, membrane and modules were prepared for use in the system. The finished modules were installed on the skid prior to shipping to the host site. The membrane and module fabrication activities are summarized as follows:

- 600 linear feet (1 m wide) each of Polaris and Proteus membrane were manufactured on MTR's commercial casting and coating lines.
- Both membrane rolls were evaluated for quality against MTR's established performance standards and were found to be in conformance.
- Four 4-inch spiral-wound Polaris modules were fabricated by MTR's manufacturing group. The same module configuration (module components and manufacturing parameters) were used for each Polaris module.
- Four 4-inch spiral-wound Proteus modules were fabricated by MTR's Research & Development group. Two configurations of the Proteus module were manufactured (different module components used). Two modules were made using configuration "A" and two modules were made using configuration "B". MTR plans to investigate the performance of A and B during the field operation.
- All modules evaluated using MTR's in-house module testing protocols were found to be in conformance with established performance standards.

In Figure 7.5, the Polaris modules are shown being installed in the test skid at Johansing Iron Works prior to shipment of the skid to the AERF.

Figure 7.5. Fabricators installing two Polaris 4-inch spiral-wound modules into the Polaris module vessel prior to shipment of the skid in May 2015 at Johansing Iron Works, Oakland, CA.

7.4.3 Site Preparation

The syngas that will be used for the test campaigns will be delivered as a slipstream from the Enerkem Alberta Biofuels plant, which is located approximately 500 ft north of the AERF. A satellite image showing the layout of the AERF and EAB facilities is shown in Figure 7.6. A 1-inch underground line from the EAB will provide syngas to AERF pilot systems on a continuous basis while the EAB is in operation. Pictures documenting the early stages of the EAB feed line construction are shown in Figure 7.7. This work and the construction of process tie-in lines at the AERF and EAB fall under EK's scope. The following summarizes the progress toward completion of the underground line and process tie-in construction:

- A kick-off meeting for the installation of the underground pipe was completed on September 16, 2015.
- Westways is the selected contractor to do the underground syngas line from EAB to AERF; the survey and welding teams started work during the week of September 21-28. This was followed by hydro-vacuum excavation work.
- The 1-inch pipe has been welded and hydro-tested.
- The pipe location was excavated and the line was installed the first week of November 2015.
- ABSA requirements were addressed by EK to provide the City of Edmonton's piping plan registration number to Westways. Westways completed the appropriate documentation, following normal protocol for ABSA, and provided it back to the City.

With completion of this work, the infrastructure for supplying syngas to the MTR skid was in place.

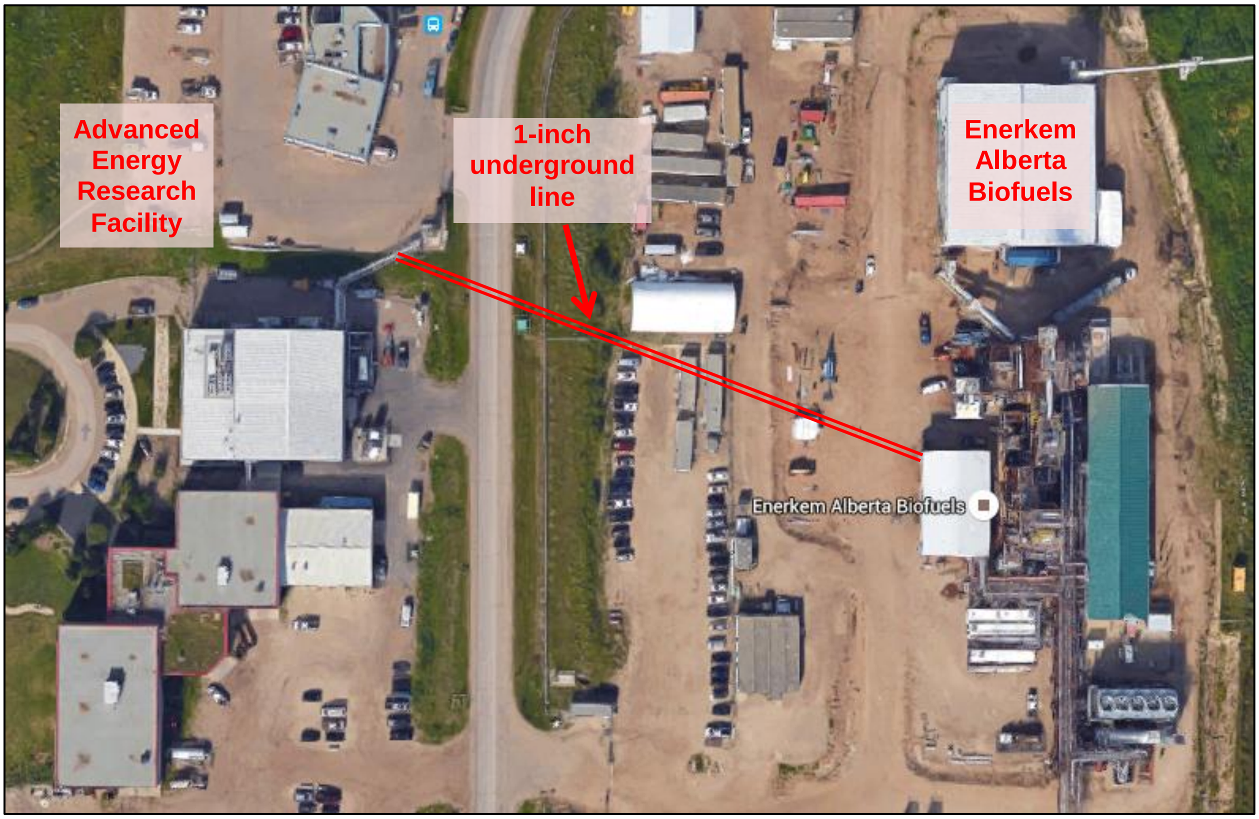

Figure 7.6. Google Maps image of the AERF and EAB showing the placement of the underground syngas feed line. Construction of the 1-inch line was completed in November 2015.

Figure 7.7. Pictures showing the early stages of EAB-AERF syngas line construction. The 1-inch pipe was welded, prepared and placed in an easy location to allow sleeve installation and drag.

In addition to preparing the site for syngas supply to the MTR skid, EK and the CoE also equipped the AERF with a GC sampling system to provide MTR with regular composition data for the six gas streams on the skid (two feed, two residues and two permeate streams). The sampling system is housed in the Class I Division 2 process area, next to the membrane skid. The GC cabinet contains a multiport selection manifold to allow multi-sampling port capability (4-port for each micro GC) from six different streams on the MTR gas separation skid, and also selected sample points in the AERF process to demonstrate membrane performance within the syngas-to-methanol process. Pictures showing the Class I Division 2 GC cabinet and the two GCs to be installed within the cabinet are shown in Figure 7.8. The following summarizes work conducted to complete installation of the GC sampling system:

- The sampling point station in the AERF facility was installed.
- The two Inficon Micro GC Fusion units and Vici Valco switching valves were installed in the process area using a CSA-approved Class I Division 2 cabinet according to the specifications.
- Electrical and communications connections to the cabinet were completed.
- Tie-in lines connecting the GC switching valves to the MTR skid were designed and installed.

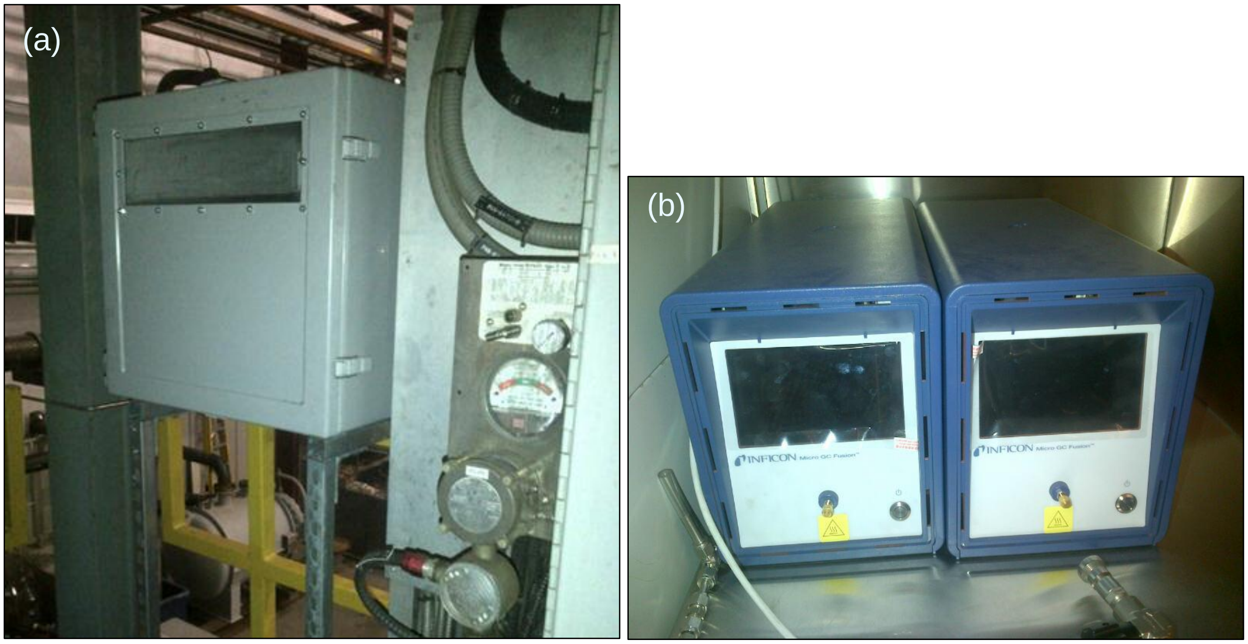

Figure 7.8. Pictures of (a) the Class I Division 2 GC cabinet and (b) the two micro-GCs installed within the cabinet.

7.4.4 Membrane Skid Installation and Commissioning

Skid installation activities at AERF are summarized as follows:

- The membrane skid was crated at the fabricator and shipped to the AERF on June 6, 2015. EK and CoE personnel placed the skid within the AERF facility on July 22, 2015.
- A local Edmonton contractor (Chuck Electrical Services) was hired by MTR to complete the tie-in of the electrical panels on the membrane skid to the utilities at the AERF. This electrical work was completed in early August 2015.
- MTR finalized the membrane skid operation manual and distributed the document to the project partners.
- Polaris membrane modules were installed in the skid at the fabricator prior to shipment to the AERF. Two Proteus modules were shipped to the AERF to be installed during pre-commissioning.

The delivered skid is shown in Figure 7.9 at the location where it is currently installed in the AERF. The skid location within the AERF test bay relative to other equipment is shown in Figure 7.10.

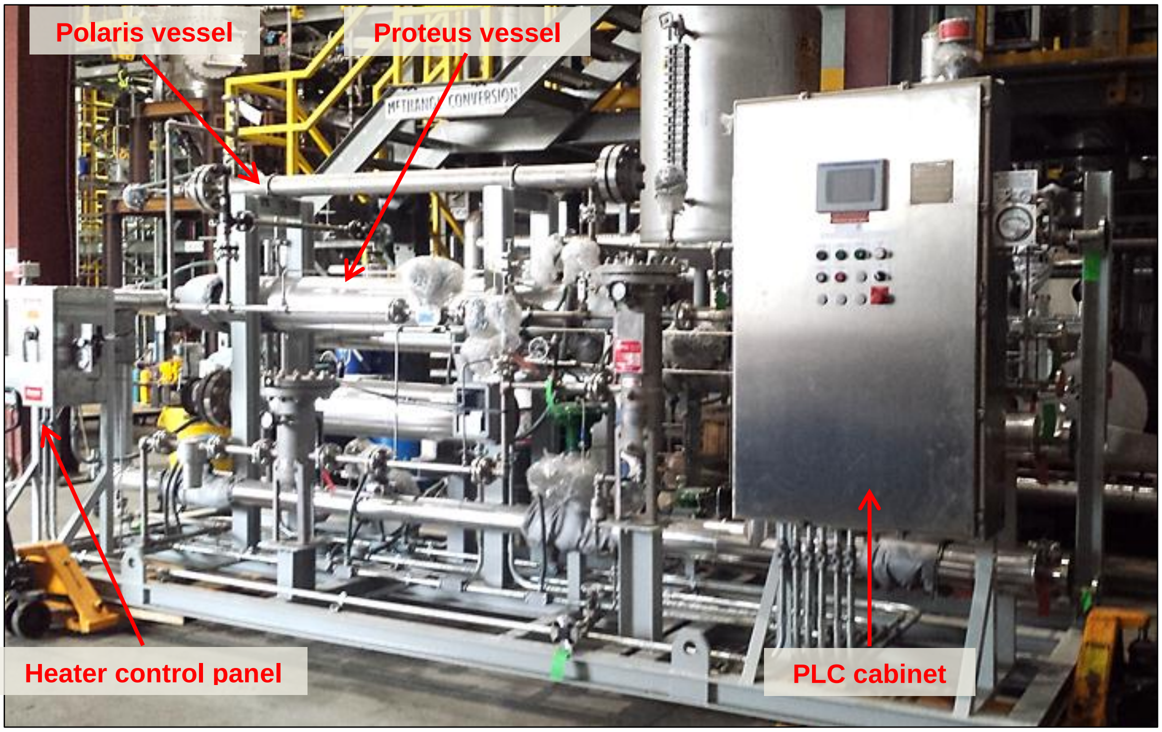

Figure 7.9. Picture of the MTR skid after it was moved to its final location in the AERF on July 22, 2015.

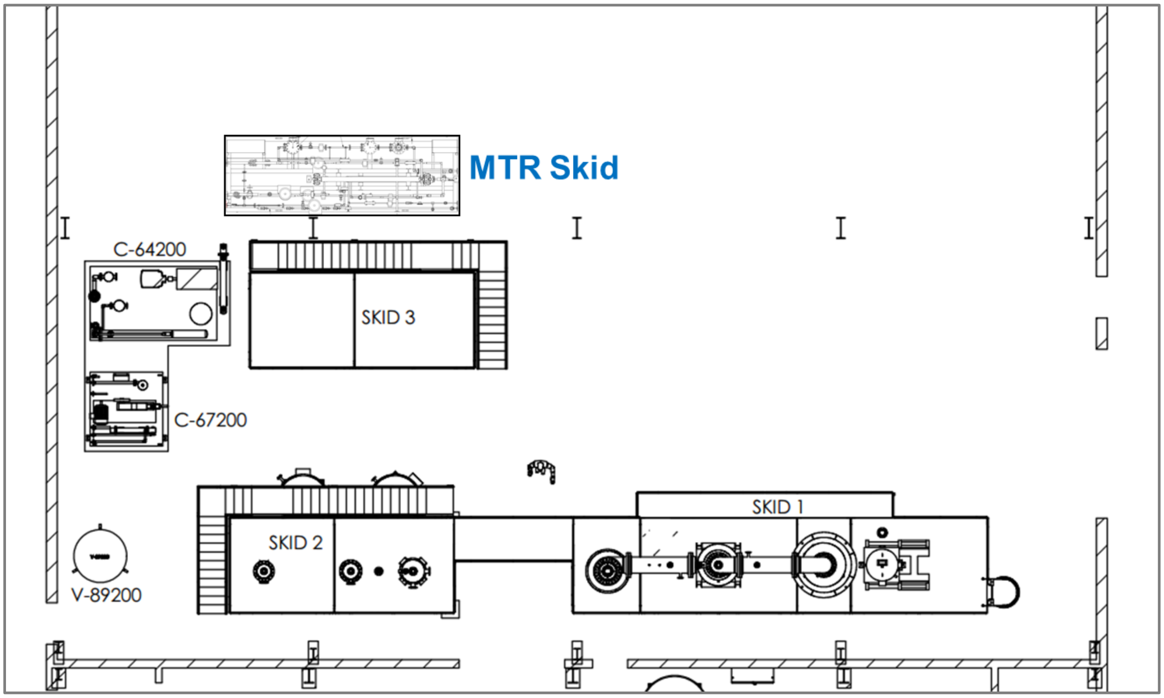

Figure 7.10. General arrangement drawing of the AERF plant floor showing the location of the MTR skid relative to the existing structures.

After installation, the following commissioning activities were completed:

- MTR personnel inspected the electrical tie-in work performed by Chuck Electric.
- CSA field certifications for the skid and the heater control panel were obtained on January 26, 2016.
- In February 2016, MTR personnel completed pre-commissioning activities, including pneumatic leak testing, PLC loop check, valve function, module installation and filter inspection.

Final tie-in of the skid and commissioning with process gas have been delayed due to operating issues at the commercial EAB facility that will be supplying syngas to the test skid. Currently, this work and the 1,000 hour field test are scheduled to occur during summer 2016. Once this field test is completed and the results are analyzed, we will issue an updated report. The following section briefly describes this future work and the field test plan.

7.5 Future Work

The last major task to be completed prior to commissioning, startup and operation is the construction of process tie-in lines at the AERF. EK currently projects that this construction will be completed in May 2016. Training, commissioning and operation will follow shortly thereafter. The following punch list summarizes the major activities planned during the next six months to complete this industrial test task:

- Complete the micro GC installation and calibration once the communication cable is put in place (May 6).
- Complete construction of process tie-in lines that connect the AERF pilot systems and the EAB syngas to the MTR skid (May 27).
- Complete operator training for EK, CoE and MTR personnel (June 10).
- Cold commission the skid with inert gas supplied by EK (June 17).
- Hot commission the membrane skid on syngas (July 1).
- Complete 1,000 hours of skid operation and parametric testing (September 1).
- Complete data analysis and process economics evaluation (September 30).
- Complete report update detailing field test findings (October 31).

The test plan for field operations at AERF is as follow:

- For the Polaris modules:
  - One module configuration, a conventional high-pressure spiral-wound design, will be tested.
  - Syngas feed pressure will be maintained at the Rectisol inlet pressure (~25 bar) for the duration of field operations.
  - Steady state performance will be collected at 3 feed temperatures: 5°C, 15°C, 25°C. We expect this parametric test to generate membrane permeance versus selectivity tradeoff data for use in future design calculations.

    - The feed flowrate will be varied at a fixed temperature. This parametric test will generate CO_2 purity vs recovery tradeoff information that will also be helpful information for design studies.
- For Proteus modules:
    - Two module configurations will be evaluated (Proteus A and Proteus B). These modules have different flow characteristics that may affect field performance. The results will be used to optimize module design.
    - The feed pressure will be maintained at the methanol reactor outlet condition (~63 bar) for duration of testing.
    - Steady state performance will be collected at 3 feed temperatures: 100°C, 120°C, 135°C. We expect this parametric test to generate membrane permeance versus selectivity tradeoff data for use in future design calculations.
    - The feed flowrate will be varied at a fixed temperature. This parametric test will generate H_2 purity vs recovery tradeoff information that will also be useful information for design studies.

This test data will help clarify the potential of Polaris and Proteus modules for use in syngas treatment.

8. WATER USE AND RECOVERY FOR A MEMBRANE CAPTURE SYSTEM

8.1 Introduction

As competition for freshwater intensifies, the efficient use of water at power stations is becoming increasingly important, particularly with regard to the consumption (unrecoverable loss) of water. In the U.S., thermal power plants are responsible for approximately 35% of freshwater withdrawals and approximately 3% of fresh-water consumption. For reference, the freshwater consumption rate for the supercritical coal-fired power plant represented in Case 11 of the DOE's Bituminous Baseline Study (BBS) is 7.7 gpm/MW_e-net.[3] When it is equipped with an amine-based CO_2 capture system, the rate of freshwater consumption is 14.1 gpm/MW_e-net, an increase of 83%.[3] In some regions, it may not be possible for an existing power plant to secure a permit for additional water withdrawal, either due to the risk of exceeding thermal discharge limits, or due to over-allocations for the watershed or basin in question.

Freshwater consumption is dominated by the needs of the power plant's cooling system. For a supercritical power plant, in the condenser, heat is transferred from steam exiting the low-pressure steam turbines to cooling water, which is then recirculated through a wet-cooling tower – see Figure 8.1. In the cooling tower, warm cooling water is contacted with air which cools the water, but results in evaporative and drift losses to the atmosphere. As cooling water evaporates, impurities in the water concentrate. To limit the overconcentration of these impurities, a portion of cooling water is discharged as a blowdown stream. Freshwater (cooling water makeup), and a smaller amount of boiler feedwater blowdown, are added to the cooling water stream to maintain a constant cooling water inventory. Cooling water makeup represents 77.5% of the total freshwater consumptive needs for the reference power plant (Case 11).[3] As freshwater comprises such a large fraction of the total water demand, it is important to examine the impacts that capture systems will have on the cooling water cycle.

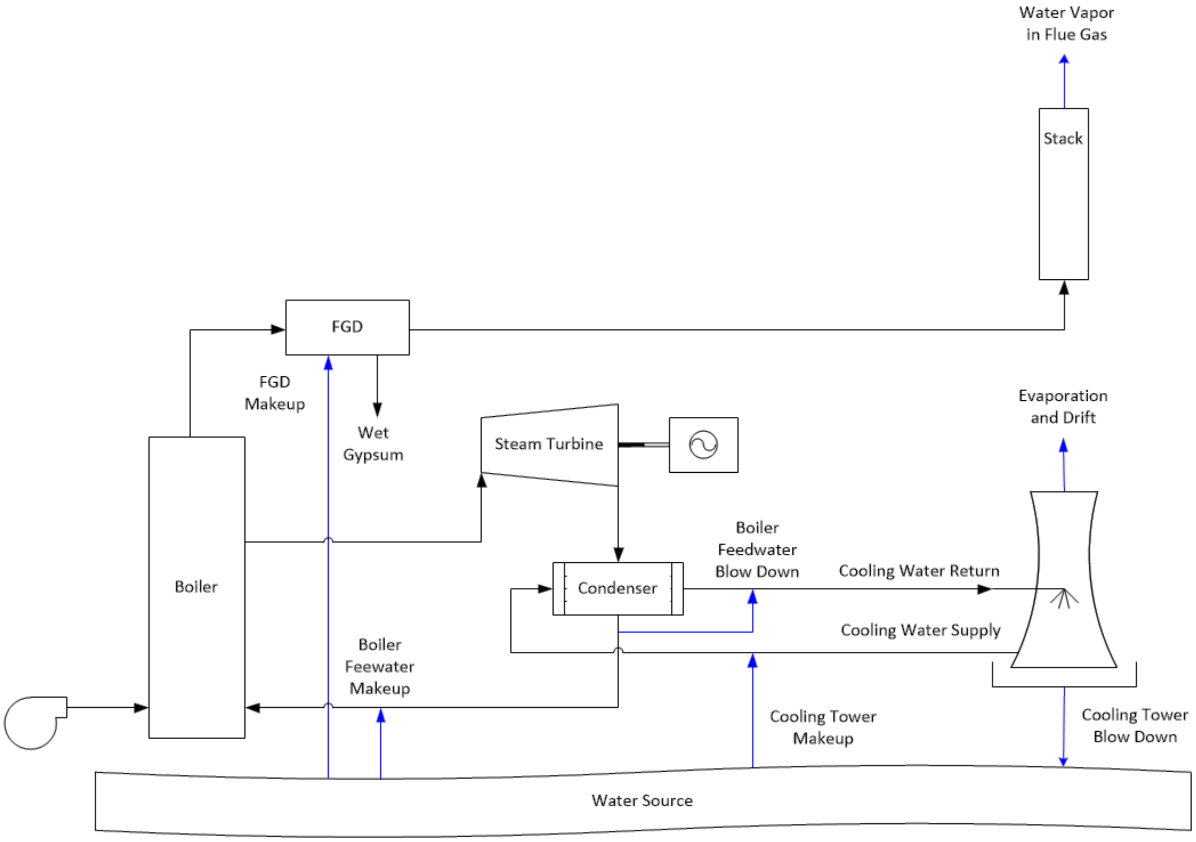

Figure 8.1. Water cycle for a supercritical power plant with a wet cooling tower.

Other freshwater demands include flue gas desulfurization (FGD) makeup and boiler feedwater makeup (internal recycle). The water requirements and flows for the Case 11 power plant are quantified in Table 8.1. Each freshwater makeup stream has purity requirements, with boiler feed water makeup being the most stringent and FGD makeup being the least stringent. The purity requirements and treatment options are discussed in detail in the EPRI water report included as Appendix B.

Table 8.1. Water Balance for Case 11.[3]

Water Use	Water Demand		Internal Recycle		Raw Water Withdrawal		Process Water Discharge		Raw Water Consumption	
	m^3/min	gpm	m^3/min	gpm	m^3/min	gpm	m^3/min	gpm	m^3/min	gpm
FGD makeup	3.6	951			3.6	951			3.6	951
Cooling tower	18.4	4,863	1.9	492	16.5	4,370	4.1	1,094	12.4	3,277
Total	22.0	5,813	1.9	492	20.1	5,321	4.1	1,094	16	4,227
Total per MW_e net		10.6		0.9		9.7		2.0		7.7

- Evaporative losses of 0.8 percent of the circulating water flow rate per 10°F of range.
- Drift losses of 0.001 percent of the circulating water flow rate.
- The blowdown rate was set to 4 cycles of concentration.

To better understand the use of water in MTR's capture system, it is helpful to review the differences between amine-based capture systems and membrane-based capture plants. All CO_2 capture systems have their own cooling water requirements, which typically include internal recycle streams both within the capture process and water exchanges with the host power plant. In DOE's baseline power plant studies, where each coal-plant case is sized to achieve a net electrical output of 550 MW_e, the gross size of the power plant must increase to compensate for the added parasitic load of the capture process. The size of the cooling water system is proportional to the amount of steam generated by the power plant, and by extension, the amount of cooling water makeup is related to the thermal input (coal feed rate) to the coal plant. For capture processes which use thermal energy to drive their separation – such as the amine-based process featured in BBS Case 12 – the cooling water and FGD makeup demand are larger than what the gross electrical output suggests. This is because a portion of the steam that would have been used to generate electricity is instead used by the capture process. Based on the gross electrical output of Case 12 (662 MW_e), the apparent cooling water demand should be 9.3 gpm/MW_e-gross, an increase of 20% over the 7.7 gpm/MW_e-net rate. Instead, the actual water consumption rate is 14.1 gpm/MW_e, an increase of 52%. This increase is mostly attributable to the capture operations, which generally include: absorber intercooling, lean solvent cooling, stripper overhead condensation, CO_2 compression intercooling, and most notably, condensing the steam used in the solvent reboiler (see Figure 8.2).

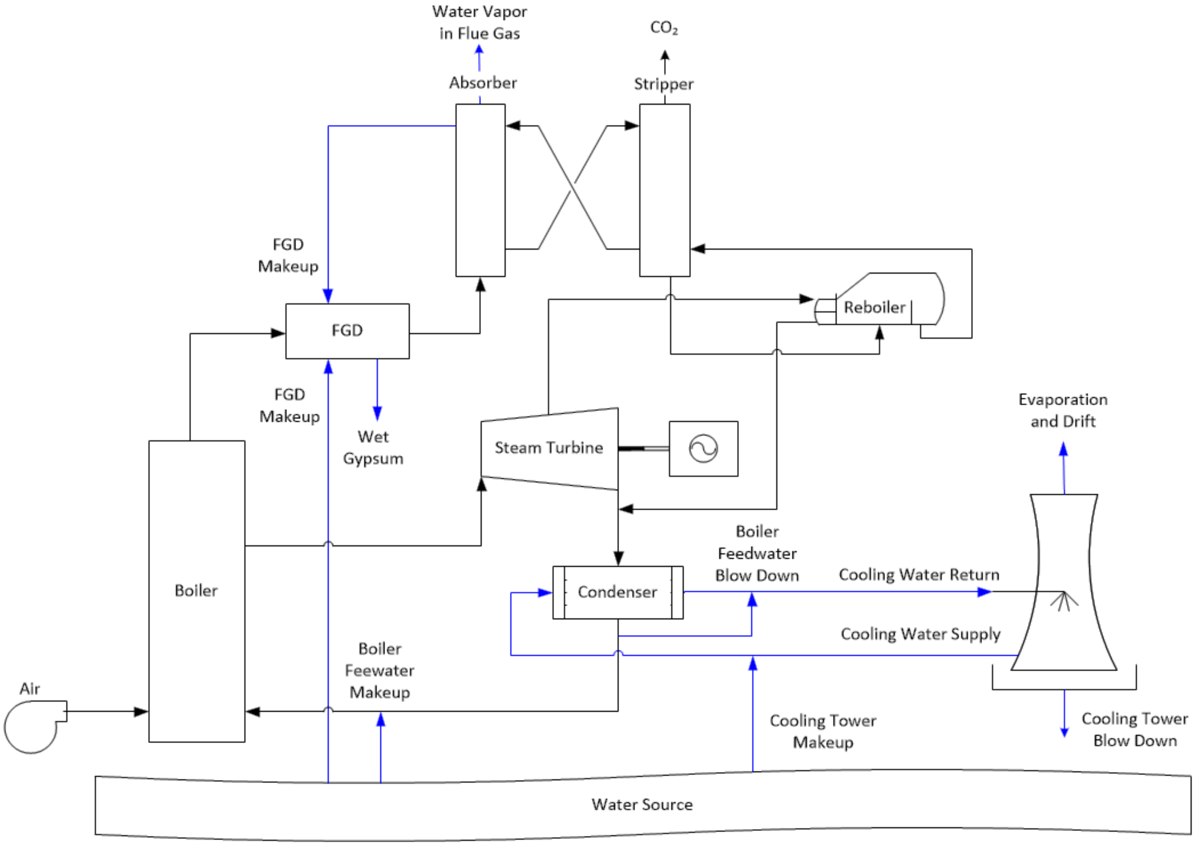

Figure 8.2. Water cycle for a supercritical coal-fired power plant equipped with an amine-based CO_2 capture system.

In contrast to Case 12, MTR's apparent cooling water demand on a gross electrical output basis is 10.9 gpm/MW$_e$-gross basis, which is close to the actual cooling water demand of 11.5 gpm/MW$_e$, an increase of only 5%. This difference between an amine-based system and a membrane-based system from the perspective of this metric is helpful for highlighting the low inherent cooling demands of membrane-based capture systems. When all cooling water withdrawals, discharges, and demands are compared, MTR's case provides meaningful reductions compared to Case 12 (see Figure 8.3). For example, the raw water consumption for the MTR system is 18% lower than the Case 12 amine system.

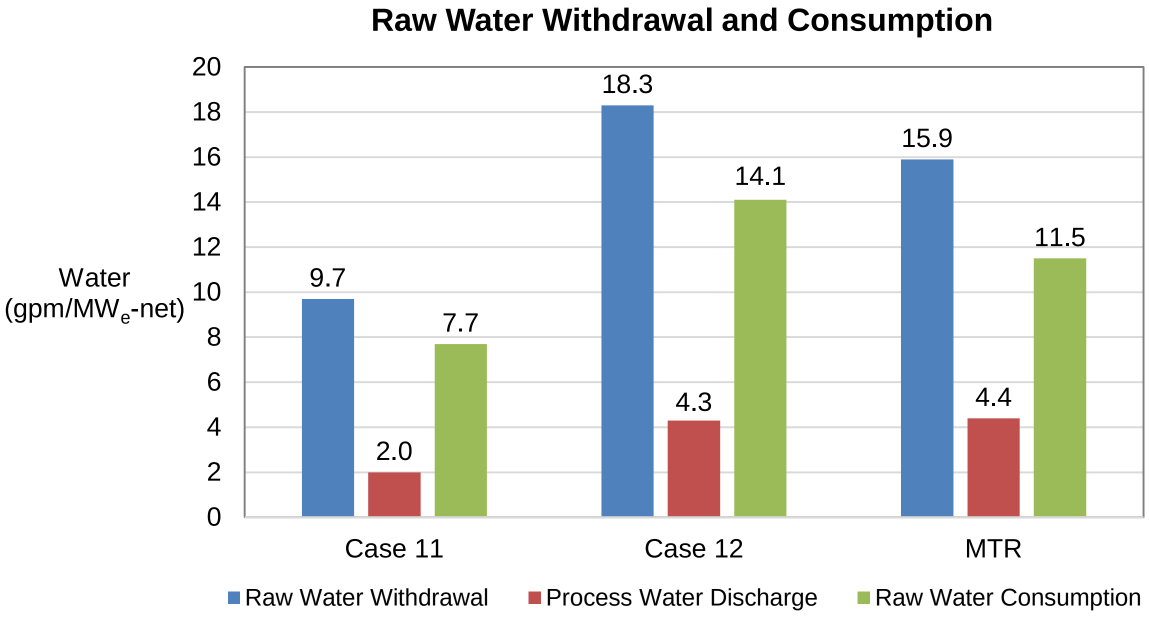

Figure 8.3. Water withdrawals and consumption for the reference no capture power plant (Case 11), a power plant outfitted with a 90% capture amine process (Case 12), and a power plant using the MTR membrane process for 90% capture.

Another difference between amine-based capture systems and MTR's capture system is the amount of water removed from flue gas in the capture step. Both capture systems require modest flue gas cooling, and both designs include a direct contact cooler (DCC) to reduce the temperature of post-FGD flue gas. As the flue gas is cooled, water condenses in the DCC. For the amine-based capture plant, the flue gas stream remains at the saturation point as it passes through the capture process and is discharged through the stack to the atmosphere. The residual water content of flue gas vented by Case 12 from the stack is 53,340 kg/hr.[3]

In MTR's capture process, the water present in flue gas permeates the membrane along with CO_2, resulting in a drier flue gas. The flue gas vented from MTR's capture process contains 7,992 kg/hr of water, which is 85% less than the water emitted in the amine-plant. This represents a 96% water capture rate for the MTR system based on post-FGD flue gas water content. The complete water cycle and water balance for MTR's system are shown in Figure 8.4 and Table 8.2.

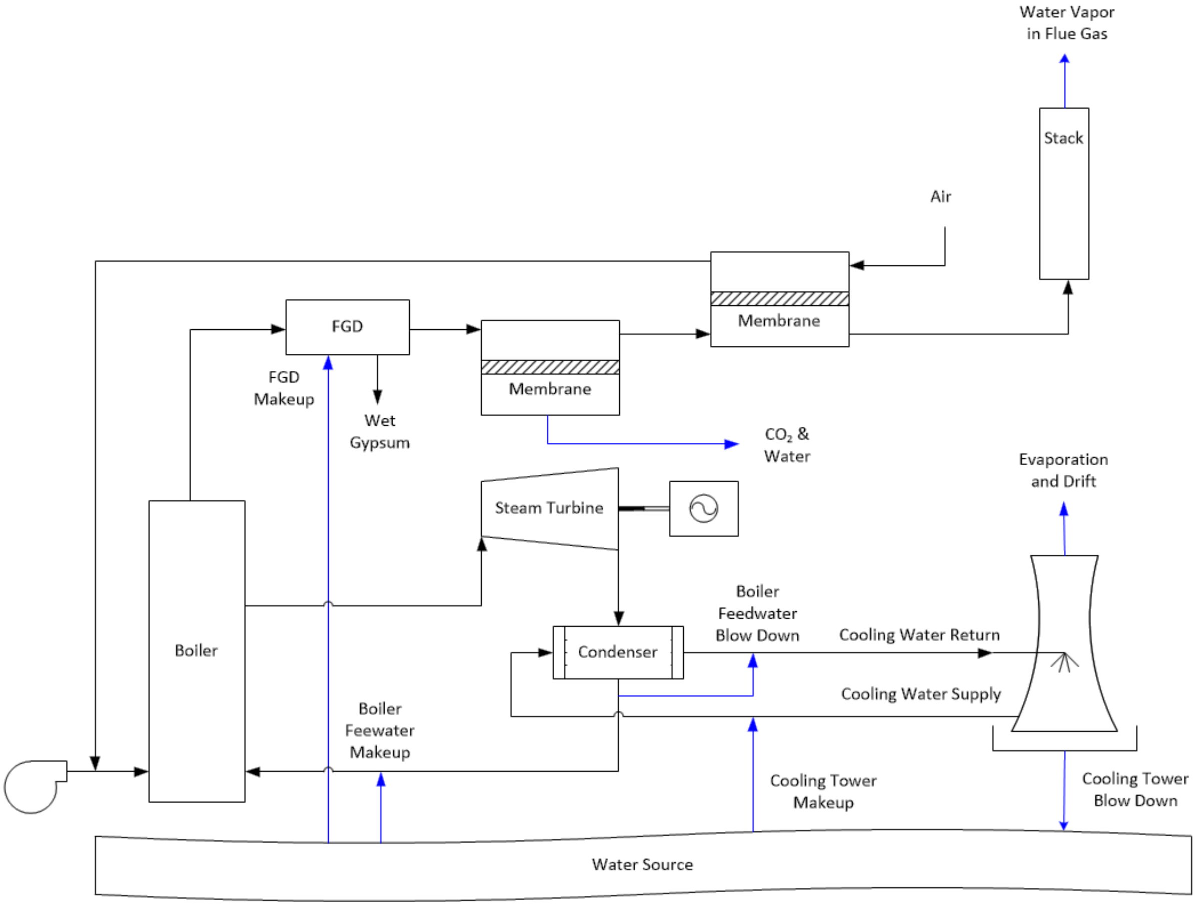

Figure 8.4. Water cycle for a supercritical coal fired power plant equipped with MTR's CO_2 capture system.

Table 8.2. Water Balance for MTR CO_2 Capture System.

Water Use	Water Demand		Internal Recycle		Raw Water Withdrawal		Process Water Discharge		Raw Water Consumption	
	m^3/min	gpm	m^3/min	gpm	m^3/min	gpm	m^3/min	gpm	m^3/min	gpm
Direct contact cooler			4	1,145	-4	-1,145			-4	-1,145
CO_2 compression and drying			1	143	-1	-143			-1	143
FGD makeup	5	1,373	3	719	2	654			2	654
Cooling tower	35	9,273		23	35	9,249	9	2,314	27	6,936
Miscellaneous	1	238		110		129		129		
Total	41	10,884	8	2,140	32	8,744	9	2,443	24	6,302
Total per MW_e net		19.8		3.9		15.9		4.4		11.5

8.2 Water Reuse Analysis

To better understand the quality of water collected by the MTR system and its potential for reuse to offset the consumptive needs described above, MTR contracted EPRI to conduct a water sampling campaign from MTR's test systems operating at NCCC. For reasons described below, the pilot plant equipment and testing platform were not conducive to providing good predictive information regarding the concentrations of contaminants in the water streams for a full-scale system, but these tests did provide useful insights into the types of species and ions that might be expected.

EPRI sampled water taken from MTR's 1 TPD and 20 TPD membrane capture systems at NCCC. The MTR test systems featured liquid ring compressors and vacuum pumps to process the post-FGS flue gas and generate the required feed and permeate pressures for the membranes. The purpose of both field test systems was to evaluate the two-step membrane performance under different gas flow conditions, including at various feed pressures. The availability of rotating equipment designed specifically for flue gas compression is non-existent. After initial unsuccessful tests with screw-type compressors, liquid ring compressors were chosen as the low-risk option because they have the ability to compress hot, wet, saturated, corrosive, and dirty gasses. By utilizing equipment known for reliable performance under demanding conditions with minimal downtime for repairs, the focus of the field tests could be entirely on the CO_2 removal performance of the membranes.

While the liquid ring compressors allowed for parametric flow and pressure testing of the MTR field test systems at NCCC, this equipment made water quality measurements difficult to interpret. For both the 1 and 20 TPD systems, the flue gas is in contact with the liquid ring of the compressor, and the gas permeating the membranes (where most of the water recovery occurs) is also in contact with the liquid ring of the vacuum pump. The water used to create the liquid ring seal inside the casing of these machines is circulated within a closed system consisting of a water supply tank and pump with minimal make-up from the cooling water supply. As a result of continuous operation, the liquid ring water becomes contaminated with particulates and other impurities found in the flue gas stream. This build-up of impurities in the liquid ring seal of the compressor and vacuum pump can then contaminate the condensed water samples taken downstream of this equipment. This effect is an artifact of the testing at NCCC, where liquid ring machines were chosen for flexibility and reliable operation. For a real-world system, a simple blower would be used to push gas through the membrane system, so there would be no opportunity for contamination of the recovered flue gas water with liquid ring seal water.

Results from EPRI water quality testing are reported in Table 8.3 and are explained in detail in Appendix B.

Table 8.3. Water Quality Test Results from MTR’s Pilot Plants During Operation at the NCCC.

Concentration (µg/mL)	20 TPD Pilot Plant		1 TPD Pilot Plant			
	Liquid Ring Compressor (Flue Gas Feed)	Post-Compression Knock-out Cooler Condensate	Compressor Inlet Knockout Condensate	Vacuum Pump Discharge Condensate	Compressor Discharge Condensate	Separator Condensate
Ammonia as N	<0.168*	0.406	0.259	0.217	0.151	0.473
Metals						
Aluminum	0.02	0.068	0.047	0.166	0.024	0.051
Arsenic	<0.005*	0.05	0.008	<0.005*	0.007	0.092
Barium	<0.005*	0.019	<0.005*	0.023	0.042	<0.005*
Calcium	0.261	10.5	<0.05	15	31	4.96
Iron	0.157	0.881	0.256	9.16	0.21	0.626
Magnesium	0.035	2.626	<0.02*	4.17	8.16	1.48
Mercury	<0.0002*	<0.0002*	0.0006	0.0002	<0.0002*	0.0021
Potassium	0.03	0.744	<0.02*	1.1	2.28	0.441
Selenium	0.011	0.609	0.068	<0.005*	0.064	0.891
Sodium	0.411	3.886	0.048	4.52	9.49	1.76
Anions						
Chloride	<0.1*	4.42	0.201	6.35	12.3	0.359
Nitrate as N	0.358	0.151	<0.02*	3.81	0.828	0.334
Sulfate	8.62	208	36.7	170	82.7	473

* Below detectable limit.

8.3 Conclusions and Future Work

This investigation provided insights into the use and possible reuse of water in an MTR capture plant, and particularly, for illustrating differences in water consumption compared to BBS Case 12. A key finding is that MTR’s capture system has a low inherent cooling demand, which can be attributed to membranes being a pressure-driven separation rather than a temperature-driven separation. This results in 18% less water consumption for an MTR membrane capture plant compared to a conventional amine capture process (BBS Case 12). Another key finding is that the Polaris membrane readily permeates water and effectively harvests water from flue gas. Compared to BBS Case 12, the MTR membrane process recovers an additional ~40,000 kg/h of water from flue gas. This recovered water may be beneficial for reuse within the plant, particularly in arid regions with water use restrictions. The EPRI water quality study was helpful in identifying the types of contaminates that may be present in recovered water, but a clear determination of the way these species are partitioned by the membrane was difficult to determine due to use of liquid ring equipment on the 1 and 20 TPD systems.

Based on these findings, MTR identified the following areas as next steps to further our understanding of power plant water issues, and to improve our ability to reuse the water harvested from flue gas.

- Identify the benefits of co-capturing water with CO_2. Recent studies have investigated the benefits of harvesting and dehydrating flue gas and using the collected water to allow a greater flexibility in siting power plants in arid locations. For example, U.S. DOE (through the *Innovative Water Reuse and Recovery* program), Southern Research Institute (in collaboration with EPRI through the *Water Research Center*), and the European Union (through the *CapWa* program). These organizations sponsored studies and R&D pilot plants using chilling/condensation, liquid desiccant, and membrane technologies. All of these programs found beneficial use for the harvested water. In this project, MTR did not reconcile the benefits (as identified by these R&D programs) of co-capturing water with CO_2. We propose to do so in follow-on work.

- Effects of dehydrated flue gas. For retrofit applications, most post-combustion capture technologies will have a cooler flue gas, compared to the stack gas from an uncontrolled power plant. Cooler stack gas results in a plume which will behave differently with regard to its exit velocity, buoyancy, mixing with ambient, and the conveyance and fate of pollutants contained within it. MTR proposes a project to study plume effects for a coal-plant (real or modeled against Case 11), utilizing MTR's CO_2 capture process for both new and existing power plant applications. As part of this study, MTR will assess the unique effects from having a relatively dry stack gas and how this will result in a plume with much less condensation (reduced visibility or not-visible).

- Dry cooling applications. For sites with limited cooling water availability, it is possible to use a variant of MTR's capture process which utilizes dry cooling. This design would also include a modified Polaris membrane with properties which allow it to work at higher temperatures. Working at higher temperatures will shift the fraction of water removed from flue gas during pre-cooling (normally in the DCC) to removal via co-capture in the membrane system. By rejecting a significant portion of process heat to the ambient air, it is possible to impound a portion of the harvested flue gas water and then use it in a hybrid wet-dry cooling system, which could be sized to have a net-zero consumption of water.

9. ENVIRONMENTAL HEALTH AND SAFETY

9.1 Introduction

MTR's capture process is generally considered to have low Environmental Health and Safety (EH&S) risk compared to other post-combustion capture technologies. This is mostly attributable to the inherent properties of the capture system, namely the passive nature in which membranes separate CO_2 from flue gas, and the simplicity of the system itself. For this EH&S assessment, MTR evaluated the environmental health and safety impacts of two operational settings. The first encompasses the membrane manufacturing process, including the production of membrane and modules. The second assesses EH&S from the perspective of the normal operation of a full-scale (550 MW_e) nth-of-a-kind (NOAK) capture plant. These assessments review specific risks, their potential impacts, and mitigation actions to protect personnel, the capture plant, and the environment.

To compare and rank different risk events, a risk assessment tool was adapted for use in this project. Risks assessment tools of this kind are used to define risk events by their likelihood of occurrence and the likely impact should they occur. Here, we assign five levels of event probability and five levels of event severity to form a risk matrix. Each of the 25 resulting outcomes were assigned to one of four risks classes (low, moderate, high, and extreme) where a low probability event with low impact is defined as a "Low Risk" event, and a high probability event with high impact is defined as an "Extreme" risk event – see Table 9.1. Each outcome can also be uniquely identified by its risk assessment code, which combines the leading letter of the *Likelihood* and *Potential Impact* (e.g. P-I indicates a *Probable* event with *Insignificant* damages).

Table 9.1. EH&S Risk Assessment Matrix.

Legend	**Damage** Likelihood	**Insignificant** No damage	**Minor** Minimal Damage	**Moderate** Moderate Damage	**Major** Major Damage	**Catastrophic** Death
🟥 **Extreme Risk** Immediate action required	**Almost Certain** 1 in 10	🟨	🟨	🟥	🟥	🟥
🟨 **High Risk** Priority action required	**Likely** 1 in 100	🟦	🟨	🟨	🟥	🟥
🟦 **Moderate Risk** Planned action required	**Probable** 1 in 1000	🟩	🟦	🟨	🟥	🟥
🟩 **Low Risk** Action by standard operating procedure	**Unlikely** 1 in 10,000	🟩	🟩	🟦	🟨	🟥
	Rare 1 in 100,000	🟩	🟩	🟦	🟨	🟨

The *Event Impacts* are defined accordingly:

- Insignificant – An event causing no injuries; OR an event causing no damage to the capture plant; OR an event causing no damage to the environment and not exceeding any discharge limits.

- Minor – An event capable of causing minor injury requiring first aid treatment; OR an event resulting in minor damage to equipment, likely causing a trip-reset-restart event or a short-duration shut-down; OR an event causing an environmental release exceeding allowed limits, but not harming the environment.

- Moderate – An event causing injury requiring hospitalization; OR an event causing equipment-level damage requiring extensive repair or replacement, resulting in a moderate-duration outage; OR an event causing an environmental release limited to the plant-site, but requiring clean-up and would require reporting.

- Major – An emergency. An event with the potential to cause serious bodily harm; OR an event causing damage to multiple pieces of equipment, requiring extensive repair and/or replacement resulting in a prolonged outage; OR an environmental release extending beyond the plant site resulting in environmental damages, requiring clean-up and likely resulting in fines and penalties.

- Catastrophic – A major emergency. An event capable of causing one or more fatalities; OR an event causing widespread damage to the capture plant possibly also damaging the host power plant; OR an event causing a major environmental release beyond the plant-site resulting in significant environmental damage/harm requiring extensive remediation, and likely resulting in exceedingly high fines and penalties.

Risk assessment studies are an integral part of process development. The steps of risk identification, mitigation assessment, and de-risking redesign, are repeated in an iterative manner throughout development to achieve a final product design and standard operating procedures that embody a low potential to harm people or the environment. The EH&S risk assessments presented below represent known risks at this stage of development. As MTR produces more and larger capture plants, we expect to identify new risks and will develop appropriate mitigation strategies to address them. However, the majority of the equipment in MTR's capture plant are common, commercial devices with significant operational experience; the likelihood of encountering large and previously unknown EH&S risks is low.

In addition to the risks assessment, MTR identified and quantified the waste streams which will be generated from the membrane production process and from the normal operation of a full-scale capture plant. Where appropriate, we comment on ways that these wastes can be minimized, or the waste streams re-used or recycled.

9.2 EH&S Analysis of Membrane Production

To support the construction of multiple full-scale capture plants, MTR will use an automated, high-volume manufacturing facility to produce both Polaris membrane and the spiral and plate-and-frame membrane elements. In this assessment, we considered a plant sized to produce two million m^2 of membrane area per year. This is the production rate required to fit two to three power plants per year. We assume that the manufacturing plant is located in the United States and that it is subject to emission limits no more lenient that what is currently required. The type of equipment used in this operation would be similar to that found in other high-volume membrane module production operations. For example, the plants run by Dow (Filmtec) and Nitto (Hydranautics) to make reverse osmosis (RO) membrane modules which use the same module configuration and similar methods of construction. These RO manufacturing plants are operated around the clock (24/7) and produce two to four million m^2 of membrane modules per year. Almost all steps in the production process are automated and robots are widely used for material handling and in production where they glue, cut and seal the membrane modules. Table 9.2 summarizes the EH&S risks associated with membrane and module production.

Table 9.2. EH&S Risks for Membrane and Module Production.

Risk	Risk Score	Mitigation Action	Comments
Personnel – general exposure to harmful chemicals	P-Mo	• Proper PPE training and equipment. • Well-defined exclusionary zones. • Monitors and alarms to detect vapor exceeding safety limits, and to detect liquid spills. • Behavioral audits to ensure compliance with safety procedures.	• Risk could be further minimized by reducing liquid hold-up and storage volumes. • Periodically check for improvements in new PPE equipment and procedures. • Periodically check for suitable replacements of “high-risk” chemicals with safer substitutes.
Uncontrolled release of solvent vapor within facility	P-Mi	• Purposeful placements of vents and fans to remove hazardous vapors at or near potential leak points. • New solvent inventory and spent solvent waste will be located in designed controlled environment.	•
Uncontrolled release of harmful vapors to the environment	P-Mi	• Abatement equipment is used to mitigate emissions to the environment.	• Proper design of abatement equipment and procedures can minimize risk of uncontrolled emissions.
Personal injury from production	L-Mi	• Use of good design practices and adherence to OSHA requirements when designing the layout of manufacturing equipment, walkways, stairs, etc. • Ample walk-around space for equipment. • Well-defined personnel exclusionary zones. • Well-lit and well-marked hazards, hot areas, pinch points, high-noise areas, low	• Potential injuries of plant personnel or visitors through normal interactions with plant equipment and structures.

Risk	Risk Score	Mitigation Action	Comments
		clearances, clearances for robotics field of movement, visual protection from automated welding machines, etc. • Ergonomics assessment for workstations, reduce repetitive motion, minimize walking by centralizing work activities. Reduce lifting requirements. • Follow best practices for manufacturing line environments (e.g. Toyota Production System) for the design of the plant, and strive for continuous improvement (Kaizen) once operational.	
Fire	R-Ma	• Standard industrial fire prevention measures including sprinkler systems, fire containment doors/walls/ceiling, hand-held fire extinguishers, and operator fire safety training.	•

The membrane and module manufacturing process generates three waste streams from the production plant:

- the water waste discharge,
- air emissions, and
- solid waste.

The water waste stream is generally biodegradable and is normally either treated in-house before discharge to local Publicly Owned Treatment Works (POTW) or directly to the POTW. This is estimated to be on the order of 0.75 gallons (2.8 liters) of wastewater per meter squared of membrane area produced.

The air emissions from membrane manufacturing facilities are typically abated. Under Best Available Control Technology (BACT) standards, emission reductions of better than 95% are achievable. Generally, vapors from organic solvents used in the manufacturing process are collected and sent to a thermal oxidizer. Abated air emissions are estimated to be approximately 8 grams per square meter of membrane area produced.

Solid waste discharge is generally limited to excess material (heals) that cannot be recovered for use with other products. In general, it is not feasible to recycle the materials used in the manufacturing process with the exception of the permeate tube. With the weight of a raw 8-inch diameter element (unfinished) estimated to be approximately 17 kg, the solid waste discharge is estimated to be approximately 80 grams per meter squared of membrane area.

Table 9.3 summarizes these high-volume membrane/module production waste streams and their treatment method.

Table 9.3. Wastes Generated From Membrane and Module Production.

Waste Component (Disposal Method)	Quantity Generated	Comments
Waste water	0.75 gallons/m^2 of membrane produced	• Waste components are biodegradable and typically discharged to the POTW.
Air emissions	8 grams/m^2 of membrane produced	• BACT is used to mitigate emissions to allowable level. • For Polaris membranes, vapors from organic solvents used in fabrication are sent to a thermal oxidizer.
Solid waste (membrane and module materials)	80 grams/m^2 of membrane produced	• Wastes from trimming and cutting of membrane in the production of membrane elements. • Wastes from discarded head-end and tail-end portions of membrane rolls. • Wastes associated with destructive QC testing. • Rejection from assembly line defects. • Non-reusable or recyclable packaging and crating.

MTR examined ways to minimize, re-use or recycle these waste streams. With regard to wastewater, the annual production of membrane will generate 1.5 million gallons of wastewater per year. One way to reduce this impact would be to use recycled water (grey water) in place of freshwater. This would have limited availability as few regions have industrial water supply systems. Another option is to partially treat the wastewater onsite with an adsorption filter bed. In this way, a portion of the freshwater consumption could be offset with a portion of the site-treated wastewater, so long as the purity of the mixture is acceptable to operations of the membrane production facility.

With regard to air emissions, for current production levels, there is no economical option to recovery or reuse the vapors which are part of the vent/exhaust streams. However, at large-scale membrane production required for a full sized capture plant, a vapor recovery system (membrane plus condensation) could be installed upstream of the thermal oxidizer to recycle organic solvents. It may also be possible to recovery and utilize waste heat generated by the thermal oxidizer. The simplest option is to use an internal recuperator to pre-heat incoming combustion air with hot exhaust gas. Also, if the vent stream has waste heat of a useful quality, heat exchangers could be installed to pre-heat water for hot water supply or to heat other lower-temperature process streams at the production facility.

The solid waste streams can be reduced through more efficient use of materials in production (i.e. better layouts to reduce cutting and trimming wastes). Also, with improved quality control and manufacturing standards, fewer materials, components, and finished membrane elements will fail Q&A tests, thus mitigating some avoidable wastes. A portion of the solid wastes streams generated during manufacturing could be reused. Here, it is helpful to be creative and imaginative to explore ways to reuse or repurpose waste streams. The easiest way to enable this is to keep like wastes together and not to comingle them. As an example, if the membrane spacer trimmings were source separated, then another party may find a useful way to repurpose them within their own products.

9.3 EH&S Analysis of a Full-Scale Capture Plant

The assessment for the full-scale plant considers all of the equipment included in the capture plant (see Appendix A for a list of equipment). The capture plant is assumed to be located immediately adjacent to the host power plant. The major interconnection points to the host plant and battery limits of the capture plant are defined to be the following streams identified in Figure 9.1:

- Stream 10 to 11: CO_2 enriched combustion air to the forced draft (FD) fan.
- Stream 26: Flue gas downstream of the FGD.
- Stream 34: Flue gas to stack.
- Stream 56: Product CO_2 (to CO_2 transportation pipeline).
- Various condensed water streams downstream of the permeate vacuum pump (returned to FGD).
- Cooling water supply and return for various cooling water heat exchangers.

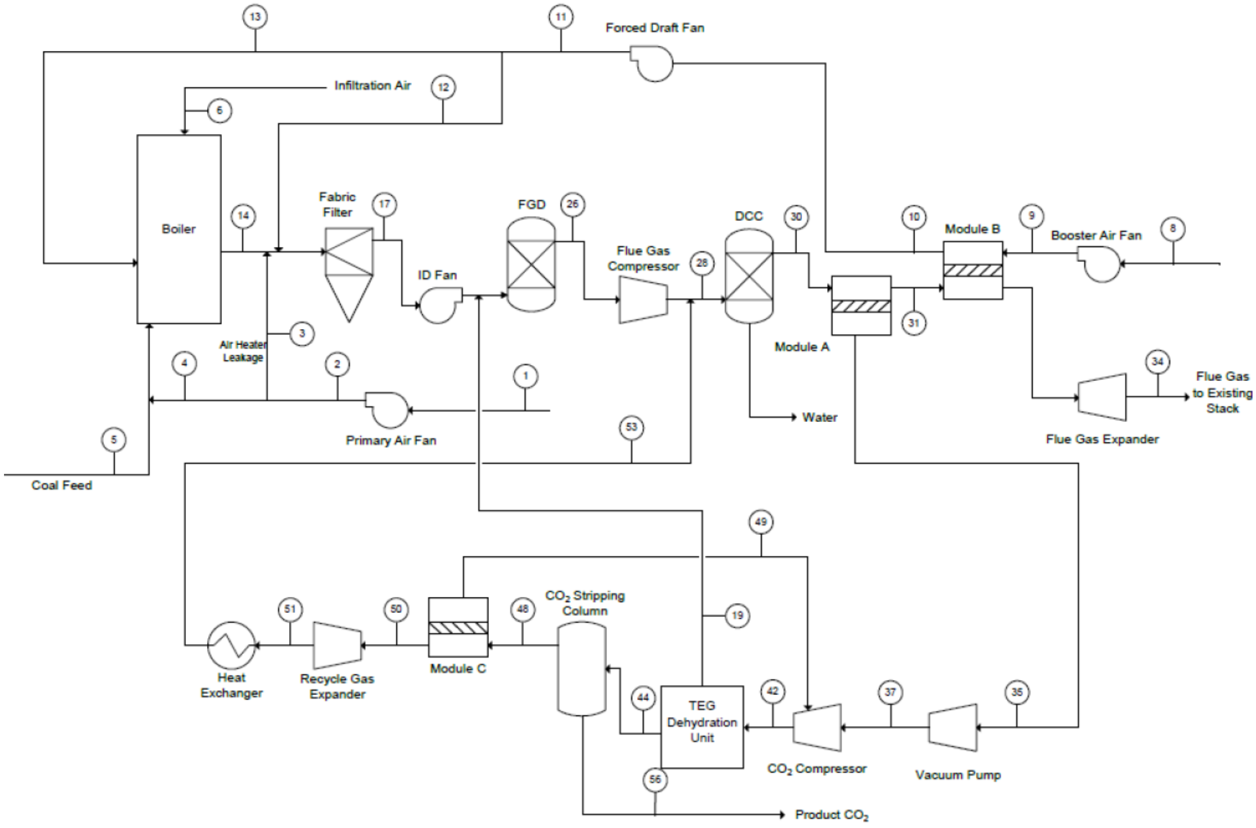

Figure 9.1. Process flow block diagram of the MTR CO_2 capture design.

The study includes all of the systems of the capture plant over the normal operation cycle, which includes all aspects of plant maintenance. Table 9.4 summarizes the risk analysis.

Table 9.4. EH&S Risks for Operation of a Full-Scale MTR Capture Plant.

Risk	Risk Score	Mitigation Action	Comments
Release of high pressure CO_2	U-Mo	• All system pressure vessels must be certified as meeting ASME standards or relevant local qualifications. • CO_2 sensors and alarms throughout the site. • Evacuation routes and muster points selected with consideration to the proximity of high-pressure CO_2 sources, areas normally downstream of them, and with consideration to low lying areas or enclosures which may trap concentrated CO_2. • Logic included in the control system to detect a sudden leak based on signals from select flow meters and pressure sensors; triggered to sound site alarms. • Check valves to isolate and limit the total volume of CO_2 that can be released in the event of a breach. • Maintenance plan includes routine inspection of high pressure sections of the plant.	• Could be a sudden, high-energy release of high pressure CO_2 from a pipe, pump or containing vessel, or a slow leak of CO_2 which could accumulate in confining, low-lying, or stagnant areas.
Lubricating oil spill	A-Mi	• All new lubricating oil shall be stored in approved containers with secondary containment in a clean, dry, and temperature-controlled environment. • Storage sites to include physical protections to protected from vehicle traffic and • Site to include appropriate inventory of oil spill cleanup kits commensurate to the volume of oil used and stored on-site.	• A spill of oil from either the on-site inventory of new oil, a collection of spent oil awaiting transport, or a leak of oil from process equipment. • Can represent a sudden leak of a large or small volume of oil (container spill), or continuous small volume leaks of oil (e.g., from a cracked gasket).
TEG leak	P-Mi	• Logic included in the control system to detect a sudden leak based on signals from select flow meters and pressure sensors. • Secondary containment near filling and draining points.	• TEG is a stable, non-toxic and non-hazardous liquid. The primary danger is contact with eyes.
Propane Refrigerant (R290) leak	U-Ma	• Sensors to detect propane leaks. • Suitable means for discharging the propane inventory.	• Propane is highly flammable. • Could be either a sudden leak and loss of propane inventory or a small leak occurring over time.

Risk	Risk Score	Mitigation Action	Comments
Rotating equipment failure	U-Mo	• Vibration sensors, temperature sensors, load sensors. Low-level and low-flow alarms for cooling fluids and lubricating fluids. • Inspect cooling and lubricating oils for metallic content (excessive wear). • Coordinate maintenance schedule for each piece of equipment to master maintenance calendar and maintenance checklists.	• The large size and rotational momentum of rotating equipment can lead to violent failures. Failure due to mechanical or electrical failure, overcurrent or undercurrent, failure of motor cooling, FOB entry into rotating equipment, detachment of blades, vanes, stators, impellors, etc.
General equipment damage resulting from improper or unsafe operation	U-Mi	• Rigorous testing of a simulated control system to test behavior of capture plant under all tripping event.	• Damages resulting from unintended operating conditions for equipment or product liquids and gases. • Could be through normal operation or
Trips, falls, general injury	P-Mi	• Use of good design practices and adherence to OSHA requirements when designing the layout of equipment, buildings, walkways, stairs, pipe-racks, etc. • Ample walk-around space for equipment • Well-defined personnel exclusionary zones. • Well-lit and well-marked hazards, hot areas, pinch points, high-noise areas, low clearances, etc. • Considerations given for routine maintenance: designated lay-down areas, easy access to fixed cranes or portable lifts and cranes.	• Potential injuries of plant personnel or visitors through the normal interactions with the plant equipment and structures.
Contaminated rain water	L-I	• Drainage plan developed with concrete pads with containment curbs, drains, sumps and collection piping appropriately sized for the site and region. • If warranted, on-site surge tanks will be installed. • Consideration given to elevating emergency or backup equipment and circuits above potential. • Consideration should be given to the potential for wastes to leak should the site flood.	• Depending on the location site, all or a portion of the capture plant could be contained in a light-frame building to provide weather protection. • System may also be enclosed, if warranted, to minimize the visual impact of the system.

The waste steams generated through normal operation of the full-scale capture plant are reported in Table 9.5. The nature of the waste streams reflects the equipment in the system.

Table 9.5. Wastes Generated During Operation of Full-Scale MTR Capture Plant.

Waste component	Amount Generated	Description
Spent membrane elements – 12" spiral (solid waste, landfilled)	105,000 kg/year	• Replace 12" spiral membrane elements at 3-year interval (at 100% CF); about 1 kg waste/m^2 of membrane replaced.
Spent membrane elements – plate-and-frame (solid waste, landfilled)	91,000 kg/year	• Replace 100 m^2 plate-and-frame membrane elements at 3-year interval (at 100% CF); about 1 kg waste/m^2 of membrane replaced.
Spent air filtration cartridges (solid waste, landfilled)	15,000 kg/year	• Pre-filters and conical primary air filter canisters replaced at 3 year intervals.
TEG (liquid, RCRA listed commercial chemical product #14163, recycled)	3,400 kg/year (7,500 L/yr)	• Some TEG will become part of the CO_2-rich stream and must be replaced. That is not reported here. The TEG waste stream assumed a full change-out of the TEG inventory once every two years.
Lubricating oil (liquid, hazardous waste if disposed of otherwise recycled)	1,000 kg/year (Estimate)	• Replacement of lubricating oils for flue gas feed compressor and motors, propane compressor and motor, expander turbine and generator, vacuum compressor and motor, pumps and pump motors. • Inventory replaced once per year.
Oil filters (hazardous waste, can be recycled)	1,000 kg/yr (Estimate)	• Oil filters are an EPA designated hazardous waste. They can be recycled by a designated recycling facility. • Prior to disposal, filters must be fully drained. There are standards for draining methods which will be followed.

Spent membrane elements represent the largest fraction of waste generated from the capture plant, totaling ~200 tonnes per year. Compared to the other listed waste streams such as filters or lubricating oils, spent membrane elements are a new type of waste, but managing the disposal is not without precedent. As previously mentioned, RO membrane elements are close proxies for MTR's Polaris membrane modules. The size, method of construction, and materials of construction are very similar. Spent RO elements are defined as a Solid Waste (not hazardous) according to EPA's Resource Conservation and Recovery Act (RCRA), 40 CFR Part § 261.2 so long as the membrane element is drained of water and is dry at the time of disposal. The standard practice is to dispose of spent RO membrane elements in municipal landfills.

With the number of reverse osmosis plants increasing, so too is the amount of waste associated with spent element disposal. In response, there are several projects investigating the option to repurpose spent RO elements as ultra-filtration membranes. In other instances, spent RO elements are mixed with other industrial wastes, such as used tires, and incinerated. MTR has not explored options to repurpose membrane elements, but it is reasonable to expect that similar options would be available as alternatives to landfilling spent Polaris modules.

Flue gas membranes will be exposed to different compounds during use as compared to RO membranes. Depending on the power plant, the types of fuel(s) used, and the environmental controls in place, flue gas can contain a wide variety of trace elements. During its time in service, each membrane module will process a large volume of flue gas during which some of these trace elements may accumulate on or within the membrane element, may permeate the membrane and become part of the CO_2 rich stream, or (most common) remain in the flue gas and be vented through the stack. Several metals known to be present in some flue gases (e.g. mercury) might trigger a RCRA special or hazardous waste designation. Should this be the case, the 12-inch elements from the vacuum membrane step (Module A in Figure 9.1) would be disposed of as hazardous waste landfill. In the U.S., there is approximately a 10:1 difference in the cost of hazardous vs. non-hazardous waste. This would increase the variable O&M costs, but considering its magnitude relative to other costs, more expensive module disposal would not materially change the cost of operations or the cost of capture.

10. PROJECT MILESTONES SUMMARY

This chapter provides a concise summary of the status of the project milestones that were defined in the Project Management Plan. Details of the work conducted to achieve these milestones have been provided in the preceding nine chapters of this report or in prior topical reports.

Table 10.1 shows the completed project Milestone Log. Each of the milestones is briefly discussed in the following list:

- The first milestone was completion of a TEA based on prior results from operation of the 1 TPD system. This analysis was summarized in a topical report titled "MTR Membrane Process for Post-Combustion CO_2 Capture: A Systems Design and Economic Analysis" that was submitted to DOE at the end of budget period 1. The primary findings from this study are summarized in Figure 1.7 and the associated text of this report.
- The second milestone was completion of the design for the 20 TPD small pilot system. This milestone was achieved at the review meeting held on February 1, 2012 attended by MTR, DOE and NCCC personnel. A presentation titled "DE-FE0005795: Pilot Testing of a 20 TPD Membrane System: DOE Design Review Meeting" was provided to DOE as documentation for this milestone. Activities associated with the design, fabrication and operation of the 20 TPD system are summarized in Chapter 4 of this report.
- The third milestone was to achieve a 50% improvement in the permeance of Polaris membranes, while maintaining a CO_2/N_2 selectivity of >30. The efforts associated with reaching this milestone performance in the form of Gen 2 Polaris membrane is described in Chapter 2 of this report. The original documentation for achieving this milestone is available in the project quarterly report from 2QFY11.
- The fourth milestone was to develop a new module design with half the pressure drop of that measured for the base case module tested at APS Cholla in the previous DOE project. This milestone was achieved during 4QFY11 and a summary of these results is provided in Chapter 2 of this report. Because of the importance of module pressure-drop, and the potential to dramatically exceed the milestone target, a separate topical report was also prepared and submitted to DOE at the end of BP1. This report is titled "Plate-and Frame Membrane Module Development for Post-Combustion CO_2 Capture". This report provides a detailed summary of the progress made in this project on module design and testing, and provides the rationale for the future DE-FE0007553 project that resulted in testing of the large plate-and-frame prototype on the 20 TPD system at NCCC.
- The fifth milestone was selection of an industrial CO_2 capture host site. Accomplishment of this milestone proved more difficult than expected. After an exhaustive search to find a domestic refinery site, we identified the Enerkem site in Edmonton as a willing host. After discussions with DOE, it was agreed to move forward with this opportunity. The summary of this work, including the preliminary industrial capture system design, is documented in the project quarterly report from 2QFY14. More details of the host site, capture process, and system design are provided in Chapter 7 of this report.
- The sixth milestone, completion of a topical report describing the impact of CO_2 recycle on boiler performance, was achieved during 1QFY14. The report titled "Effect of CO_2-Enriched Air on Combustion Performance: Pilot-Scale Evaluation" was previously

submitted to DOE. A summary of findings from this work is discussed in Chapter 5 of this report.

- The seventh project milestone was to cut the cost of module production from the initial \$500/m^2 to half this value, and to show a pathway to the ultimate goal of \$50/m^2. This milestone was achieved and is described in Chapter 2, as well as the project quarterly report from 3QFY13.
- The eighth project milestone was completion of fabrication and installation of the 20 TPD system. As described in Chapter 4 of this report, the system fabrication was completed in 2QFY14, and after delivery to NCCC in April 2014, installation was completed in 4Q2014. These accomplishments were originally documented in project quarterly reports, as well as in MTR presentations at the DOE CO_2 Capture Technology Meetings of 2014 and 2015.
- The ninth project milestone was completion of small pilot system testing. As described in Chapter 4 of this report, the small pilot system operated during two campaigns at NCCC (PO-2 and PO-3) during the first six months of 2015. Operation was completed on June 30, 2015, and was originally documented in the 3QFY15 project quarterly report.
- The tenth project milestone was to complete testing of the 1 TPD system at NCCC. Originally, this milestone included testing of a CO_2 liquefaction step added to the 1 TPD system. The objective of this proposed liquefaction study was to examine the CO_2 purification portion of the MTR process design shown at the bottom of Figure 1.4. However, after discussions with DOE and NCCC, it was agreed to omit this liquefaction testing for two primary reasons:
    - In a separate DOE project (DE-FE0006138) that started and ended during this program, MTR designed and successfully operated a CO_2 liquefaction system at NCCC's pre-combustion test facility. This membrane system produced high purity CO_2 from coal-derived syngas containing about 10% CO_2 in bulk nitrogen. Because of the similarity of this stream to the feed to the 1 TPD post-combustion system, the liquefaction membrane system design was identical for the two cases. With this new information from the DE-FE0006138 project, the plan to test liquefaction on the 1 TPD system became redundant.
    - MTR recommended and DOE agreed that it was more important to continue operation of the 1 TPD system to collect important membrane lifetime data rather than to interrupt operation of this system for liquefaction modifications. At the same time, MTR could focus resources on the commissioning and operation of the 20 TPD system, which was ongoing at this time.

  For these reasons, the tenth milestone was modified to be completion of 1 TPD system operation, which was accomplished in 4QFY15 and described in Chapter 3 of this report.
- The eleventh project milestone was completion of a topical report describing operation of the industrial field test system. As described in Chapter 7 of this report, while the industrial capture system was designed, fabricated, and installed at the host site, it has not yet operated due to delays associated with supply of feed gas to the system. This delay is outside of our control and completely dependent on the host site, who is committed to completing the field test. Once the field test is completed, MTR will prepare and issue a topical report to DOE describing the findings.

- The final project milestone was completion of this final report, including the updated TEA (discussed in Chapter 6 and Appendix A). An additional component added to this milestone was completion of a Case Study of the MTR capture process applied to a ~20 MW_e power plant by ISTC. This report titled, "Case Studies on the Potential Retrofit of Two Illinois Power Plants with MTR Membrane Technology for Post-Combustion Capture and Assessment of the Impacts of Recycling CO_2 for Combustion when using MTR Capture Technology on Downstream APCDs," was submitted to DOE in final form in 2QFY16. Among the key findings of this report was that removal of CO_2 from flue gas will impact the calculation of air pollutant emissions by conventional Continuous Emissions Monitoring Systems (CEMS) installed at existing power plants.

Table 10.1 Project Milestone Log

Project Milestone Description	Project Duration - 60 months Start: 10/1/10 End: 9/30/15										Planned Start Date	Planned End Date	Actual Start Date	Actual End Date	Comments (notes, explanation of deviation from baseline plan)
	Budget Period 1		Budget Period 2				Budget Period 3								
	Year 1		Year 2		Year 3		Year 4		Year 5						
1. Complete the techno-economic report for a 550 MW_e power plant membrane system using 1 TPD (0.05 MW_e) system data	xxx	xxX									1QFY11	4QFY11	1QFY11	2QFY12	Completed in 2QFY12
2. Complete design of the 20 TPD (1 MW_e) system	xxx	xxX									1QFY11	4QFY11	1QFY11	2QFY12	Completed in 2QFY12
3. Show at least a 50% improvement in membrane CO_2 permeance with CO_2/N_2>30 at standard conditions	xxx	xxX									1QFY11	4QFY11	1QFY11	2QFY11	Performance achieved by 2QFY11 already met project targets
4. Demonstrate a new module design with half the pressure drop of the baseline module	xxx	xxxX									1QFY11	4QFY11	1QFY11	4QFY11	Performance achieved on schedule during 4QFY11
5. Complete selection of an industrial CO_2 capture test site and begin design of this system.			xxx	xxx							3QFY12	3QFY13	1QFY13	2QFY14	Completed during 2QFY14
6. Complete a topical report describing the impact of CO_2 recycle on boiler performance			xxx	xxX							3QFY12	3QFY13	2QFY13	1QFY14	Completed in 1QFY14
7. Reduce cost of module fabrication to \$250/m^2 and show pathway to \$50/m^2			xxx	xxx	xxx	xxX					1QFY12	4QFY13	1QFY12	3QFY13	Completed ahead of schedule

Project Milestone Description	Project Duration - 60 months Start: 10/1/10 End: 9/30/15										Planned Start Date	Planned End Date	Actual Start Date	Actual End Date	Comments (notes, explanation of deviation from baseline plan)
	Budget Period 1		Budget Period 2				Budget Period 3								
	Year 1		Year 2		Year 3		Year 4		Year 5						
8. Complete fabrication and installation of the 20 TPD system			xxx	xxx	xxx	xxX					1QFY12	4QFY13	2QFY13	4QFY14	Fabrication completed in 2QFY14; installation completed in July 2014
9. Complete testing with the 20 TPD system							xxx	xxx	xxx	xxX	1QFY14	4QFY15	4QFY14	3QFY15	Completed as of June 30, 2015
10. Complete testing with the 1 TPD system at NCCC*							xxx	xxX			1QFY14	4QFY15	4QFY14	4QFY15	Completed July 17, 2015
11. Complete a topical report describing industrial CO_2 capture field test results							xxx	xxX			1QFY14	4QFY15	4QFY14	See comments in text	Test system installed; awaiting syngas supply
12. Complete LCOE analysis, write Final Report									xxx	xxX	1QFY15	4QFY15	3QFY15	2QFY2016	This report

* With DOE approval, this milestone was modified to omit CO_2 liquefaction testing as described in the text.

11. CONCLUSIONS, LESSONS LEARNED, AND RECOMMENDATIONS

The primary goal of this DOE-funded project was to scale-up the MTR membrane CO_2 capture process to the small pilot test stage through the design, fabrication, and operation of a 20 TPD test unit. This objective was achieved. The 20 TPD small pilot system successfully completed over 1,000 hours of operation at NCCC, consistently capturing >85% of the CO_2 in a flue gas slipstream. In addition to this achievement, other project accomplishments included laboratory membrane and module optimization at MTR, validation field testing of new membrane generations on a 1 TPD membrane system at NCCC, completion of boiler modeling and testing at B&W to quantify the impact of recycled CO_2 on boiler performance, and an updated TEA by EPRI/WP that clarifies the potential of the membrane process. Together, these efforts have advanced the MTR membrane capture technology through small pilot testing (TRL 6), reducing the risk associated with future scale-up.

Specific project accomplishments, and when appropriate the project milestones they met, are as follows:

- A Gen-1 Polaris membrane (CO_2 permeance = 1,000 gpu) accumulated over 10,000 hours of stable operation on flue gas using a bench-scale 1 TPD system at NCCC, providing valuable membrane lifetime information. This testing greatly reduced the risk associated with uncertainly in the durability of CO_2 capture membranes when treating coal-derived flue gas.
- A Gen-2 Polaris membrane (CO_2 permeance = 1,700 gpu) with performance exceeding project targets (CO_2 permeance >1,500 gpu with no loss in selectivity) was scaled-up to commercial roll-to-roll production and validation tested on the 1 TPD system at NCCC. An advanced Polaris membrane (CO_2 permeance = 3,000 gpu) was developed at lab scale, showing a pathway to performance needed to meet DOE targets.
- Membrane module production costs were cut by 60% to $\$200/m^2$, meeting project cost reduction targets. A pathway to achieving commercialization cost goals ($\$50/m^2$) was developed.
- A 20 TPD small pilot system was designed, built and installed on schedule and budget. The system showed obvious size and simplicity advantages over similar capacity solvent units also installed at the NCCC host site. The membrane system operated during two NCCC campaigns in 2015 and collected over 1,000 hours of operation with flue gas, achieving 90% CO_2 capture in parametric testing and consistently capturing over 85% of the flue gas CO_2.
- A newly-designed plate-and-frame sweep module was tested at NCCC on the 20 TPD system and demonstrated to have >4 times lower sweep-side pressure drop compared to spiral sweep modules. This improvement will save ~10 MW_e of blower energy at full scale.
- CFD modeling and pilot boiler testing at B&W confirmed the viability of selective CO_2 recycle sweep operation. The impact of recycled CO_2 on boiler efficiency was quantified (0.75% efficiency loss) and incorporated into the project TEA. At this modest efficiency

loss, the selective recycle process has a significant net positive effect on capture costs, particularly at high capture rates (>70%).

- The project TEA shows that the capture costs at 90% capture for the MTR membrane process with selective recycle ($52/tonne) are lower than the baseline solvent capture approach ($56.5/tonne). A sensitivity study indicates that the membrane process shows a minimum capture cost around $35/tonne at partial capture rates of 40-60%. At lower capture rates (<40%), corresponding to proposed EPA Clean Power Plant regulations, the difference in costs between the 2-step selective recycle membrane design and a simple 1-step process without recycle is small. For these cases, it may be preferable to use the simple 1-step process because it doesn't have any impact on boiler operations.
- An industrial capture host site was identified at a waste-to-biofuels facility near Edmonton, AB. A 1 TPD test system for evaluating advanced CO_2-selective and-H_2-selective membranes was designed, built, and installed. Operation has been delayed until late 2016 due to issues with the facility providing the feed gas slipstream.
- A water consumption analysis indicates that the MTR membrane process uses about 18% less water than the base case MEA process. In addition, the vent gas from a membrane system contains 85% less water than the flue gas vent from the MEA case, and this additional recovered water may find beneficial reuse within the plant.

Based on the experience gathered in this project, we believe membranes have a role to play in CO_2 capture. At 90% capture, the MTR membrane process shows a small cost advantage over the base-case amine system even without full n^{th}-of-a-kind optimizations. Moreover, at lower capture rates, membrane capture costs are significantly reduced and they are likely to show significant advantages over solvent systems (which show relatively constant capture costs as a function of capture rate). This is especially relevant given currently proposed CO_2 emissions limits, which correspond to partial capture rates of <40% from an average coal-fired power plant. In addition, membranes enjoy a number of advantages over other capture approaches that are not readily incorporated into capture cost estimates. For example, membrane systems are inherently environmentally friendly and involve no hazardous chemical handling, disposal, or emissions issues, in contrast to many solvent systems. Depending on the power plant location, such environmental considerations may be determinative in selection of a CO_2 emissions mitigation approach. Likewise, the water consumption advantage enjoyed by membranes, and described in Chapter 8, could have a significant influence on capture technology choice in arid regions with strict water use limitations.

Over the course of this 5-year research program, there were many lessons learned. A summary of some of the important ones follows:

- Development of a new technology for a large-scale application like CO_2 capture from power facilities takes substantial time and resources. As illustrated in Figure ES1, it took about 8 years and >$20 million in DOE funding (plus >$4 million in MTR resources) to advance the MTR membrane capture technology from concept (TRL 2/3) through small pilot testing (TRL 6). MTR considers this development to have proceeded relatively successfully without major setbacks.
- It is important to be flexible enough with the technology development pathway to accommodate, in parallel with scale-up activities, early stage developments that could have a significant impact on the process economics. An example that occurred in this

program was identification of a new module design that dramatically lowers pressure drop, and the subsequent DE-FE0007553 project to build a prototype that was then incorporated into the small pilot testing. Allowing this parallel development, alongside the 20 TPD scale up, will yield large energy savings (as much as 10 MW_e) at full scale. Similarly, advanced manufacturing concepts, although themselves at perhaps early technology readiness levels, have a potential to significantly lower costs (and even improve performance), and are worthwhile to pursue alongside testing scale-up.

- Testing at bench scale (1 TPD) with real flue gas is critical before moving to small pilot testing. This is particularly true for second generation capture approaches like membranes, where balance of plant components are different than for solvent systems and typically have not been operated with flue gas before. The experience we gained by operating the 1 TPD membrane system for long periods of time at NCCC was key to selecting robust components (rotating equipment, valves, etc) and designing proper upset contingencies for the 20 TPD unit. This preparation allowed the 20 TPD system to run very smoothly at NCCC, taking full advantage of the experimental time and ensuring no cost over runs.
- The availability of a host site like NCCC is invaluable for technology developers. Compared to prior field tests at other sites, this project benefited significantly from the infrastructure and focused expertise available at NCCC.

Despite the significant progress made in this program, there is still work to do to minimize risks and make membrane-based CO_2 capture a viable commercial option in the near future. We recommend the following steps:

- At the next stage of scale-up (for example, 10 MW), it will be useful to operate the membrane system over a range of capture rates, including those consistent with expected EPA regulations. Essentially all of the industrial inquiries that we receive on CO_2 capture ask about either the lowest capture cost regardless of capture rate or capture systems to meet proposed EPA regulations. Testing at these partial capture conditions will allow for optimization and refined cost estimates, which is particularly important for technologies like membranes, where system design and costs can change significantly depending on CO_2 removal requirements. Moreover, although B&W have shown that selective recycle to the boiler is feasible, this approach still introduces additional risk for the MTR process that can be removed at lower capture rates by using a single-stage system.
- As described in this report, substantial progress has been made in improving membrane performance and reducing costs. However, to meet nth plant targets, additional membrane and module manufacturing improvements will be required. It will be important to continue these optimizations during the next stage of scale-up.
- In addition to membrane improvements, there is potential to reduce overall capture costs through balance of plant optimizations. For example, most membrane – as well as some sorbent – capture approaches utilize vacuum pumps to achieve CO_2 separation. These will be very large vacuum machines making up a significant fraction of the cost and energy use of the capture plant. However, because there is no near-term market or funding available, there is no organized effort to improve vacuum efficiency and costs. This is in contrast to CO_2 compression, where substantial research funding has been devoted to optimizing this compression equipment. Better vacuum pumps would be beneficial to membranes and other capture technologies.

- The potential benefits of co-capture of flue gas water along with CO_2 by the membrane process should be examined further. Recent studies have investigated the benefits of harvesting and dehydrating flue gas and using the collected water to allow a greater flexibility in siting power plants in arid locations. For example, U.S. DOE (through the *Innovative Water Reuse and Recovery* program), Southern Research Institute (in collaboration with EPRI through the *Water Research Center*), and the European Union (through the *CapWa* program). These organizations sponsored studies and R&D pilot plants using chilling/condensation, liquid desiccant, and membrane technologies. All of these programs found beneficial use for the harvested water.
- While much of the research focus on post-combustion CO_2 capture has been on coal-fired power generation, North America has seen a dramatic shift to natural gas-based power production in the past few years. This trend seems likely to continue, and eventually capture from natural gas power plants will be required to meet CO_2 emission reduction obligations. MTR has described a selective exhaust gas recycle (SEGR) approach for membrane-based capture from natural gas that is analogous to the coal process described in this report. This SEGR concept has been well-received in literature, but not yet tested because funding is required to get turbine manufacturers to participate in an evaluation of the process. We believe a feasibility study to flesh out this SEGR concept as either an all membrane approach or as a hybrid capture system would be worthwhile.

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APPENDICES

Appendix A

MTR CO_2 Capture Process for a Supercritical Coal-Fired Plant: Technical and Cost Evaluation Final Report

Appendix B

Reuse Assessment of Water Recovered from MTR's Post-Combustion, Membrane-Based Process

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EPRI, INC.

MTR CO_2 CAPTURE PROCESS FOR A SUPERCRITICAL COAL-FIRED PLANT

Technical and Cost Evaluation
Final Report

prepared for:

EPRI | ELECTRIC POWER RESEARCH INSTITUTE

and

MTR Membrane Technology & Research

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SYNOPSIS

This report documents the technical and cost evaluation the MTR CO_2 capture process for a supercritical coal-fired plant.

Disclaimer

This study report is conceptual in nature and represents the work of WorleyParsons performed to the recognized engineering principles and practices appropriate for conceptual engineering work and the terms of reference provided by WorleyParsons' contractual Customer, EPRI, Inc. (the "Customer"). This study report may not be relied upon for detailed implementation or any other purpose not specifically identified within this study report. The extent of any warranty or guarantee of this study report or the information contained therein in favor of the Customer is limited to the warranty or guarantee, if any, contained in the contract between the Customer and WorleyParsons.

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PROJECT 108103-07885 - TECHNICAL AND COST EVALUATION

REV	DESCRIPTION	ORIG	REVIEW	WORLEY-PARSONS APPROVAL	DATE	CLIENT APPROVAL	DATE
A	Draft for Review	EM E Mace	HG H Goldstein	David B. Stauffer D Stauffer	12 Jun 2015		
B	Final Report	EM E Mace	HG H Goldstein	David B. Stauffer D Stauffer	30 Jun 2015		
C	Removed copyright, modified Disclaimer	EM E Mace	HG H Goldstein	David B. Stauffer D Stauffer	08 Jun 2016		

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Appendices

LIST OF EXHIBITS

Prepared by WorleyParsons Group
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George Booras, EPRI Principal Technical Leader

Joseph Swisher, EPRI Project Engineer/Scientist

Tim Merkel, MTR Principle Investigator

Brice Freeman, MTR Consulting Engineer

Acronyms and Abbreviations

acfm	Actual cubic feet per minute
AR	As received
atm	Atmosphere (14.696 psi)
B&W	Babcock and Wilcox
BBS	Bituminous Baseline Study
BEC	Bare erected cost
BFD	Block flow diagram
BOP	Balance of Plant
Btu	British thermal unit
Btu/h	British thermal units per hour
Btu/kWh	British thermal units per kilowatt hour
Btu/lb	British thermal units per pound
CC	Carbon Capture
cf	Cubic feet
CF	Capacity factor
CFM	Cubic feet per minute
CO	Carbon monoxide
CO_2	Carbon dioxide
CS	Carbon steel
DCC	Direct contact cooler
DOE	Department of Energy
EPA	Environmental Protection Agency
EPRI	Electric Power Research Institute
FD	Forced draft
FGD	Flue gas desulfurization
ft	Foot, Feet
FO&M	Fixed operations and maintenance
gal	Gallon
GJ	Gigajoule
GHG	Greenhouse gas
GPM	Gallons per minute
h, hr	Hour
H_2O	Water
H&MB	Heat and mass balance
HHV	Higher heating value
hp	Horsepower
HP	High pressure
Hz	Hertz
ID	Induced draft
in. Hg	Inch mercury
in. W.G.	Inch water guage
IP	Intermediate pressure
kV	Kilovolt
kW, kWe	Kilowatt electric
kWh	Kilowatt-hour
kWt	Kilowatt thermal
lb	Pound
lbm	Pounds, mass
lb/h	Pounds per hour
lb/MMBtu	Pounds per million British thermal units
lbmol	Pound mole
lbmol/h	Pound moles per hour
lb/MWh	Pounds per megawatt hour
LP	Low pressure
MEA	Monoethanolamine (CO_2 scrubber solvent)
MM	Million
MMBtu	Million British thermal units
mole%	Mole percent (percent by mole)
MTR	Membrane Technology Research
MVA	Mega volt-amps
MW,MWe	Megawatt electric
MWh	Megawatt-hour
MWt	Megawatt thermal
N_2	Nitrogen
N_2O	Nitrous oxide
N/A	Not applicable
NETL	National Energy Technology Laboratory
NO_X	Oxides of nitrogen
NRE	Non-recurring engineering (costs)
O_2	Oxygen
O&M	Operation and maintenance
OFA	Over fired air
PC	Pulverized coal
PF	Power factor
PFD	Process flow diagram
PM	Particulate matter
ppm	Parts per million
ppmv	Parts per million volume
ppmvd	Parts per million volume, dry
psi	Pounds per square inch

psia	Pound per square inch absolute
psid	Pound per square inch differential
psig	Pound per square inch gage
Qty	Quantity
SA	Secondary air
scf	Standard cubic feet
scfd	Standard cubic feet per day
scfm	Standard cubic feet per minute
scmh	Standard cubic meter per hour
SCR	Selective Catalytic Reduction (NOx control)
SO_2	Sulfur dioxide
SO_X	Oxides of sulfur
SS	Stainless steel
STG	Steam turbine generator
TEG	Tri-Ethylene Glycol
mt, tonne	Metric ton (1,000 kg)
TPC	Total plant cost
tph	Tons per hour
U.S.	United States
V	Voltage
VO&M	Variable operations and maintenance
V-L	Vapor liquid portion of stream (excluding solids)
vol%	Volume percent
wg	Water gauge
wt%	Weight percent
°C	Degrees Celsius
°F	Degrees Fahrenheit

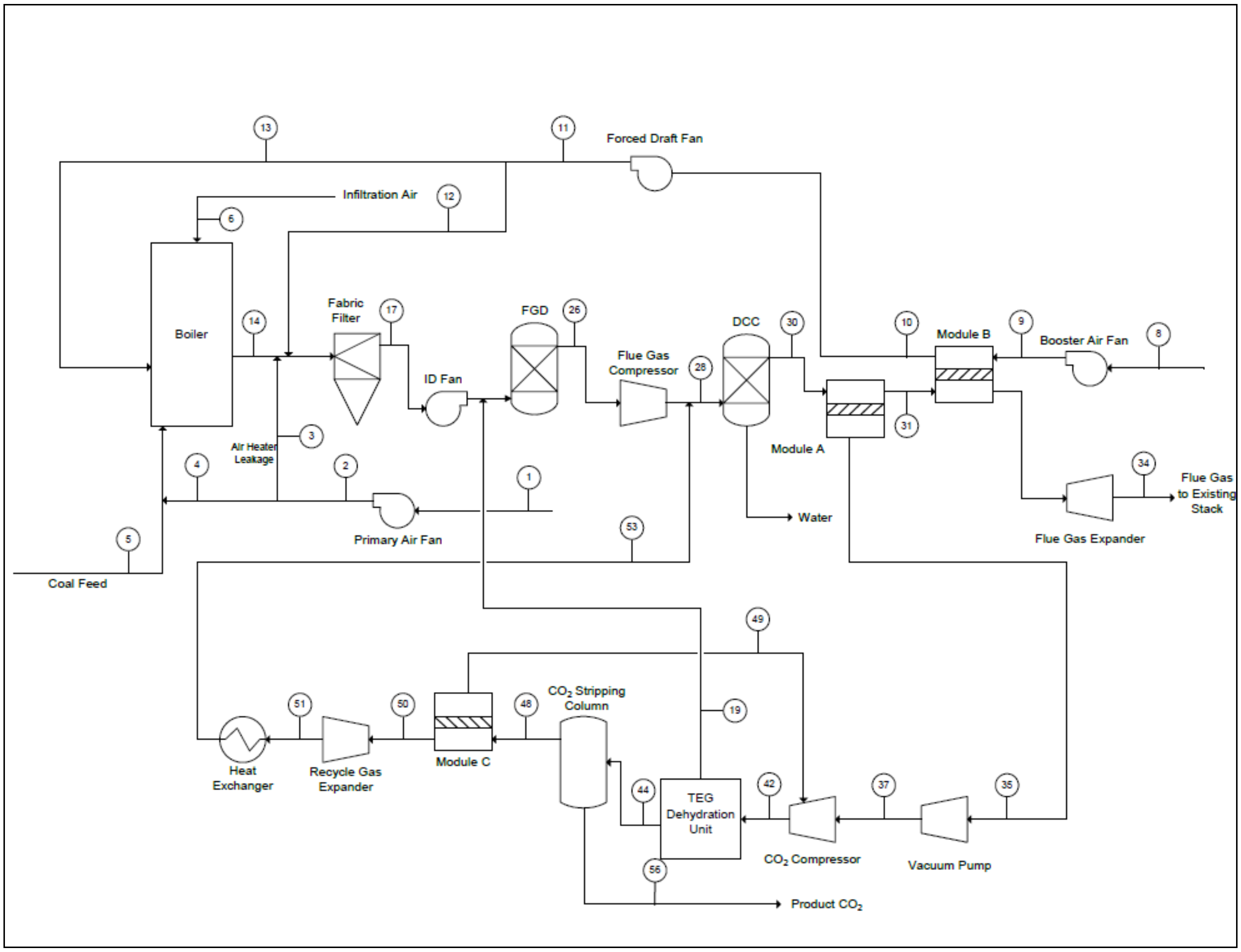
1. EXECUTIVE SUMMARY

The objective of this evaluation by WorleyParsons is to provide technical and cost input to EPRI for the economic evaluation of MTR's CO_2 membrane capture system in a greenfield application to a reference pulverized-coal (PC) plant. The chosen reference PC power plant design is Case 11 of the "Bituminous Baseline Study" (BBS) for post-combustion capture technologies [1]. The nominal net plant output for this study is set to 550 MWe. The plant performance and cost is compared to Case 11 (supercritical pulverized coal) and Case 12 (supercritical pulverized coal with MEA CO_2 Capture) from the BBS report, with updated costs based on the National Energy Technology Laboratory's (NETL) "Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases" [2].

The BBS supercritical PC reference plant is a 3500 psig class Rankine cycle with nominal steam turbine throttle characteristics of 3,500 psig/ 1,100°F /1,100°F. The plant emission control is based on low-NO_X burners (LNB) w/ OFA and SCR for NO_X control, fabric filter for PM control and wet FGD producing waste grade gypsum for SO_2 control. The Case 11 reference plant Net Plant HHV efficiency is 39.3%.

MTR's CO_2 membrane capture system centers around the use of CO_2 permeable membranes in the flue gas stream downstream of the wet FGD. A simplified block flow diagram is presented in Exhibit 1-1. A flue gas compressor is added to facilitate operation of the membranes at 2 bar to minimize the required membrane area and to improve the CO_2 separation. CO_2 that is separated by the first membrane modules, which are cross flow modules, is sent for compression and additional purification. The CO_2 specification of less than 10 ppm O_2 in the CO_2 product requires the introduction of a purification column. A small stream rejected by the compression / purification system is recycled back to the CO_2 separation system. Flue gas retentate leaving the first CO_2 module passes through the counter flow modules where fresh air is used to sweep the permeate side of the membrane. The sweep air / permeate leaving the counter flow membrane is integrated into the secondary air system of the existing PC steam generator.

Exhibit 1-1: CO_2 Membrane Process Block Flow Diagram



Note: Modules A & C consist of cross flow membranes. Module B consists of counter flow sweep membranes.

Compared to the base case secondary air, the sweep air fed to the secondary air system is vitiated of oxygen (circa 18 mol% O_2), has an increased CO_2 level (circa 8 mol% CO_2) and increased mass and volumetric flows. These changes to the secondary air lower the boiler efficiency by approximately 2 %. The largest performance effect of the MTR CO_2 membrane system on BBS Case 11 is the significantly increased auxiliary load associated with the required turbo-machinery. Additional discussion of the effects of the CO_2 system utilization on the plant are presented in Section 4.

A summary of the cases compared in this study is presented in Exhibit 1-2.

Exhibit 1-2: Evaluation Matrix

Case	Description	CO_2 Capture/ Technology	CO_2 Capture/ Compression	Notes
BBS Case-11	Supercritical PC (Base Case)	None	None	3,500/1,100/1,100
BBS Case-12	Supercritical PC (Ref Case w/ CC)	MEA (Amine) Absorber	90% capture/ 2,215 psia	Reference Amine CC case
MTR	Supercritical PC (MTR Case w/ CC)	CO_2 Membrane	90% capture/ 2,215 psia	Focus of this Evaluation.

The technical and cost information developed for the evaluation of the CO_2 membrane retrofit are based on a conceptual level of detail.

1.1 Performance Summary

The plant configuration for the MTR Case is the same as BBS Case 11 with the exception that the MTR CO_2 membrane separation, purification and compression systems have been added for CO_2 capture. The nominal net output is maintained at 550 MWe by increasing the boiler size and turbine/generator size to account for the greater auxiliary load imposed by the CO_2 capture systems.

A performance summary for the three cases listed above is presented in Exhibit 1-3. The MTR CO_2 membrane capture system is fully capable of capturing the targeted 90% of the CO_2 generated. The impact to the generation unit is such that the gross generation increased by approximately 200,395 kW to 780,795 kW, while the auxiliary load increased by approximately 196,893 kW to 227,303 kW, yielding a net generation of 553,492 kW, which is slightly over the target of 550,000 kW. Compared to the BBS Case 12 reference case net plant efficiency of 28.4%, the MTR efficiency of 28.6% is slightly improved.

Exhibit 1-3: Summary of Technical Performance and Cost Results

	BBS Case 11	BBS Case 12	BBS Case 11 with MTR Capture
Gross Power Output, kWe	580,400	662,800	780,795
Auxiliary Power Requirements, kWe (BOP)	30,410	47,340	40,796
Auxiliary Power Requirements, kWe (CC)	0	65,490	186,507
Auxiliary Power Requirements, kWe, (Subtotal)	30,410	112,830	227,303
Net Power Output, kWe	549,990	549,970	553,492
HHV Thermal Input, kWth (MMBtu/h)	1,400,162 (4,778)	1,934,519 (6,601)	1,934,964 (6,602)
Net Plant HHV Efficiency, %	39.3	28.4	28.6
Net Plant HHV Efficiency Penalty, % Point	0.0	10.9	10.7
Boiler Efficiency (%, HHV)	88.0	88.0	86.2
Raw Water Withdrawal, m^3/MWh Net (gpm/MWh net)	2.2 (9.7)	4.2 (18.3)	3.6 (15.8)
Raw Water Consumption, m^3/MWh Net (gpm/MWh net)	1.7 (7.7)	3.2 (14.1)	2.6 (11.4)
CO_2 Generated, kg/h (lb/h)^A	440,322 (970,728)	608,090 (1,340,587)	608,438 (1,341,353)
Capture Efficiency, %	0.0	90.2	89.9
CO_2 Emitted, kg/h (lb/h)	440,322 (970,728)	59,697 (131,608)	61,239 (135,007)
CO_2 Emissions, kg/MWh Gross (lb/MWh Gross)	759 (1,673)	90 (199)	78 (173)
CO_2 Emissions, kg/MWh Net (lb/MWh Net)	801 (1,765)	109 (239)	111 (244)

Notes:

A. Based on 100% carbon conversion.

1.2 Cost Estimating Summary

The capital cost estimates developed herein have an accuracy level of -15%/+30%, consistent with the conceptual level of the study and the BBS reference cases. The results of the capital and O&M cost estimation effort are represented in Exhibit 1-4. Per Exhibit 1-4 the Specific Total Plant Cost (TPC) of the MTR case is approximately 75% higher than the Base Case (Case 11), compared to approximately 80% higher for BBS Case 12.

Exhibit 1-4:
Comparative Summary of Capital and Operating Costs (June, 2011USD, $x1000)

Item	BBS Case 11	BBS Case 12	MTR Case
Capital Costs			
Total Plant Costs (TPC)	$1,089,771	$1,959,399	$1,915,457
Specific TPC, ($/kw, net)	$1,981	$3,563	$3,461
% increase in Specific TPC	Base	79.9%	74.7%
Annual Operating Costs			
Fixed Costs	$38,829	$64,138	$62,858
Variable Costs	$31,688	$54,089	$46,634
Fuel Costs	$104,591	$144,504	$144,542

1.3 Alternate Configuration / Application

Current proposed EPA regulations (40CFR60, Subpart Da dated January 8, 2014) regarding CO_2 emissions imply an approximate 40% capture rate from the base case emissions. Operating at this reduced capture rate would simplify the MTR system and provide significant performance and cost benefits relative to 90% capture. This concept is discussed in Section 8.

2. INTRODUCTION

Membrane-based CO_2 capture processes are an alternative to solvent-based absorption processes for large-scale separation of CO_2 from power plant flue gas. Membrane Technology and Research, Inc. (MTR), based in Newark, CA, has developed membranes and a membrane-based process for capturing 90% of the CO_2 emitted by a coal-fired power plant. These membranes and process elements have been tested in DOE-sponsored tests at the bench and small-pilot scale. The Electric Power Research Institute (EPRI) is performing a techno-economic analysis of a full-scale design of MTR's process as part of a US Department of Energy (DOE)-sponsored project.

The primary objective of this study is to assist EPRI in performing an economic evaluation of MTR's CO_2 capture process by providing performance and cost information based on membrane performance data.

This document presents the technical and cost evaluation the MTR process for a supercritical coal-fired plant with MTR's CO_2 capture process operating at 90% overall capture of CO_2. The power plant design is based on Case 11 of the "Bituminous Baseline Study" (BBS) for post-combustion capture technologies [1]. The nominal net output of the plant with capture is 550 MWe. The plant performance and cost is compared to Case 11 (supercritical pulverized coal) and Case 12 (supercritical pulverized coal with MEA CO_2 Capture) in the BBS report, with updated costs based on the National Energy Technology Laboratory's (NETL) "Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases" [2].

Current EPA regulations regarding CO_2 emissions imply an approximate 40% capture rate. Operating at this reduced capture rate would simplify the MTR system and provide significant performance and cost benefits relative to 90% capture. This concept is discussed in Section 8.

The evaluation scope includes:

1. Developing an evaluation basis that defines essential technical and functional requirements for establishing a conceptual design based on a nominal 550 MW (net), greenfield PC plant for the post combustion capture technologies, identical to that used for Case 11, super-critical pulverized coal (PC) with CO_2 capture.
2. Technical evaluation of the power plant design scenario with the MTR CO_2 membrane, including:
    A. Heat and Mass Balances (H&MBs)
        i Air, flue gas and membrane system (BFD, PFD & H&MB table)
        ii Supercritical steam cycle (H&MB diagram)
    B. Performance Tables
        i Gross and net power generation, and itemized auxiliary loads
        ii CO_2 rich stream summary information
        iii Plant air emissions
        iv Carbon, sulfur and water balances

        v Overall energy balance

    C. Engineering & Documentation

        i System description

        ii Qualitative analysis of the anticipated impact of the MTR system on the new plant design (particularly impact of air sweep on the boiler performance and systems)

        iii Major equipment list

3. Cost estimates (target accuracy -15%/+30% for known items with additional contingency for novel/ less certain items).

    A. Capital cost estimate. (Factored estimates for traditional / known items). Reference case costs should be escalated to mid-2011 basis.

    B. O&M cost estimate, including estimated membrane life and replacement cost as provided by MTR

3. EVALUATION BASIS

This section contains a summary of essential technical and functional requirements that are used as a basis in establishing conceptual designs for this study.

The following design basis parameters in this report are assumed to be identical to the BBS report:

- Plant location and ambient conditions
- Coal composition and heating value
- Sorbent characteristics
- Environmental requirements
- Flue gas desulfurization (FGD) and selective catalytic reduction (SCR) system design and performance
- Steam cycle configuration, design and performance
- Cooling system configuration, design and performance
- Balance of plant systems configuration, design and performance
- Equipment redundancy requirements
- Plant capacity factor

Performance and heat and mass balances are performed on the same basis as in the BBS report, with a nominal net power output of 550 MWe, which corresponds to 100% steam throttle flow rate, and referenced to the ambient conditions defined in the BBS report.

3.1 Modeling Approach

A critical input for determining the impact of the CO_2 membrane application for separating CO_2 from the flue gas of a coal fired plant is the development of a heat and mass balance (H&MB) and corresponding performance estimate. The study methodology includes performing steady-state simulations of the various power plant technological islands with process simulation models. To this end, several different specialized computer modeling software programs were employed, each with its own niche in the overall analysis. The modeling software is listed below, followed by a brief description of how it was utilized within the analysis.

3.1.1 MTR Membrane Performance

MTR provided the performance for the three different membranes utilized in the evaluation [3]. Performance information was provided in the form of an H&MB table and corresponding PFD. The information provided by MTR was utilized by WorleyParsons in the supplemental analyses.

3.1.2 Aspen Plus

WorleyParsons utilized Aspen Plus software to evaluate the impact of the membrane integration on the boiler, air and flue gas gaseous streams. The Aspen Plus analysis is complicated by the presence of three recycle streams: the sweep air from the counter-flow membrane module (Module B), the CO_2 purification system recycle stream and the TEG dehydration unit vent recycle. The presence of the recycle streams required that WorleyParsons and MTR iterate between their modeling software to ensure sufficient convergence of the results. Since the majority of the membrane application impact is to the unit's gas side, the Aspen Plus analysis represents the heart of the overall analysis. The Aspen Plus model was initially calibrated using the H&MB data for Case 11 of the BBS report.

3.1.3 Boiler Performance

WorleyParsons utilized the boiler performance inputs based on a yet-to-be-publicized December 2013 report prepared by Babcock & Wilcox for MTR. This input was used to set the boiler oxygen levels as recommended by B&W, as well as to estimate the boiler efficiency loss after incorporating the membrane sweep air. The increased CO_2 flowing through the boiler is a result of the vitiated air from counter-current membrane module feeding the forced draft fans in lieu of fresh air.

3.1.4 GateCycle – Steam Cycle Performance

WorleyParsons utilized General Electric's GateCycle software to address the impacts to the supercritical Rankine steam cycle resulting from integration of the membrane system. Initially, the GateCycle model was calibrated using the H&MB data for Case 11 detailed in the BBS report.

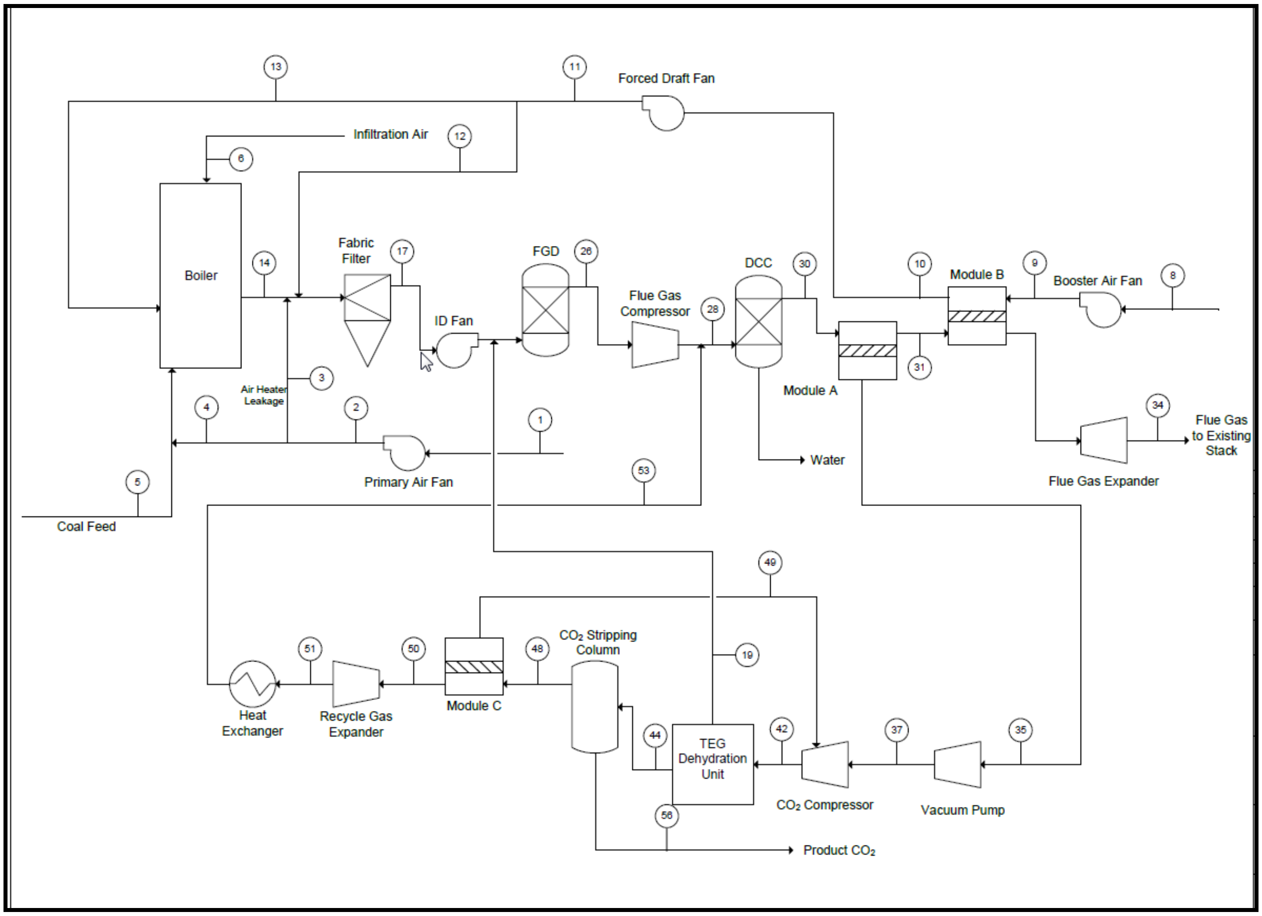
3.2 MTR CO_2 Membrane Requirements and Performance

The objective of the MTR CO_2 membrane capture system is to capture 90% of the CO_2 from the flue gas of a supercritical pulverized coal (PC) plant while minimizing the auxiliary load through implementation of the MTR cross-flow and counter-flow modules employing permeate vacuum and sweep air respectively. The heart of the CO_2 membrane process is the cross-flow module (Module A) and the counter-flow sweep module (Module B), both of which operate on a partial pressure driving force. The cross-flow membrane achieves the CO_2 partial pressure driving force through a pressure gradient achieved by balancing feed compression and a permeate vacuum.

The counter-current sweep module achieves its CO_2 partial pressure driving force through the use of an air sweep stream. The advantage of using the air sweep stream is that the process air leaving the module can be utilized in the PC boiler. The entrained CO_2 is then recycled back into the process, thus increasing the overall capture rate[A]. A block flow diagram (BFD) of this process is presented in Exhibit 3-1

[A] Where less than 90% CO_2 capture is required, a membrane process that eliminates the counter-current sweep module could be entertained, thus eliminating the performance impact on the steam generator and secondary air system. If such a capture system were desired, this has the potential to be a significant sweet spot for the membrane system. This is discussed further in

Exhibit 3-1: Block Flow Diagram of MTR CO_2 Capture Process



The preliminary design requirements for the MTR CO_2 membrane requirements are presented in Exhibit 3-2.

Exhibit 3-2: MTR CO_2 Membrane Inlet Requirements [4]

Criteria	**Limit/ Target**	**Note**
Particulates	Not yet known	The particulate matter (PM) is the greatest concern of all of the anticipated contaminants, as it can lead to life-ending fouling/ clogging. As such, the PM limit will be discussed in a dedicated subsection below.
SO_X	N/A	The membrane is robust with respect to SO_X. Both SO_2 and SO_3 are polar and will permeate through the membrane.
NO_X	N/A	The membrane is robust with respect to NO_X. No limit was specified.

Section 8.

Criteria	Limit/ Target	Note
O_2	N/A	There is no limit on the O_2 in the flue gas stream.
Feed Temperature	<70°C	The membrane is robust at temperatures below 70°C. The preferred range is 10-50°C for better membrane performance.
Superheat Temperature	N/A	Flue gas can be saturated with water when fed into the first membrane step. Since the membrane is very permeable to water, the water content in the feed decreases rapidly. As a result, water condensation on the feed side of the membrane appears unlikely. The concern with liquid condensation is that it would block flow channels causing undesirable pressure drop.
Feed Pressure	2.0 bara	Analysis by MTR [5] indicates that the membrane area is reduced most appreciably by being between 2 and 3 bara, while the plant's net power output will be notably higher at 2 bara versus 3 bara. As such, the 2 bara pressure will form the basis of this analysis.
Flow Rate	Design Flow	For lower capture rates, membrane bypass can be used. However, since 90% capture is the target, a bypass is not envisioned.
Heavy Metals	N/A	MTR membrane is not adversely affected by heavy metals.

3.2.1 Particulate Matter Membrane Requirement

Particulate matter is a potential membrane contaminant of much interest to the project members as deposition of the PM may lead to fouling and clogging of the membrane. Unfortunately there is great uncertainty regarding how much PM will deposit in the membrane versus simply pass through the membrane. High-efficiency candle filters could be added, but they add a significant pressure drop, require substantial real estate and are costly. On top of that, the candle filters may not be necessary. Similarly, a wet ESP could also be added, but the added cost and complication of handling the wet ash is likely unnecessary.

The membrane flow path is measured on the order of a millimeter. Although no information is presently available, the particulate size leaving the fabric filter / FGD is postulated to be on the order of several microns to submicron. Since 1,000 microns fits between a 1 mm flow path, it seems possible that much of the ash could be carried thorough the entire membrane. It is also postulated that additional PM removal will occur in the direct contact cooler (DCC) downstream of the FGD. . To date, bench-scale testing with post-FGD coal flue gas at the National Carbon Capture Center (NCCC) over a period of 10,000 hours has shown no significant impact of residual particulates on membrane system performance. Therefore, this first phase of evaluating the membrane integration application will be based on the assumption that high efficiency candle filters downstream of the fabric filter / FGD are not needed. [6]

3.2.2 MTR CO_2 Membrane Performance Parameters

The CO_2 membrane performance data was extracted from simulation data provided by MTR and incorporated into the Aspen Plus H&MB developed for the project. General performance observations regarding the membrane are presented in the following paragraphs.

The MTR CO_2 membrane is based on the Polaris™ membrane which allows polar molecules (e.g., H_2O, CO_2, SO_2, SO_3, H_2S, NO_2) to permeate. Although SO_3 is expected to permeate through the CO_2 membrane into the CO_2 product, being a hydrophilic molecule it would end up with the water removed by the CO_2 compression process.

Oxygen gas, O_2, a non-polar molecule, will preferentially be rejected by the membrane, and depending upon the feed concentration, could comprise up to approximately 1.5 mole % of the CO_2 product stream leaving the cross-flow membrane (Module A).

The O_2 content of the sweep air leaving the counter-flow membrane module is approximately 18% O_2. That is, some O_2 in the incoming air diffuses through to the flue gas side due to an O_2 partial pressure difference, thus somewhat depleting the sweep air of O_2, while the CO_2 that permeates into the sweep air also acts to dilute the O_2 concentration. The air sweep flow can be limited to about 50% (and higher) of the total combustion air and maintain near-maximum benefits of the sweep (i.e., increased CO_2 driving force). [7] Nevertheless, for a base case, the project team has chosen to supply all of the PC boiler's secondary air from the counter-flow module, corresponding to roughly 80 wt% of the total combustion air.

3.2.3 MTR CO_2 Membrane Design Parameters

The module vessels that MTR currently uses for natural gas CO_2 removal are 26 ft (7.9 m) long x 5 ft (1.5 m) diameter cylinders. These high pressure vessels weigh 15 tons (13.6 tonnes) fully loaded with 2,600 m^2 (28,000 ft^2) of membrane. Adapting this technology for low-pressure flue gas, MTR estimates a weight of 7 tons (6.4 tonnes) including skid supports. Ultimately, MTR will look at redesigning module vessels for flue gas. One design being considered now is rectangular modules where a 1 m x 1 m x 1 m box would contain 1,000 m^2 (10,800 ft^2) and weigh less than 2 tons (1.8 tonnes). These rectangular modules could be easily stacked to increase the packing density and thereby reduce the footprint. [8] This information is summarized in Exhibit 3-3. The required area by module is presented in Exhibit 3-4.

Exhibit 3-3: Module Vessel Design Parameters [8]

Module	Area	Dimensions	Weight (Fully Loaded)	Notes
Current Module Vessels for NG CO_2 removal	2,600 m^2 (28,000 ft^2)	26 ft x 5 ft (7.9 m x 1.5 m) L x D	High P. App.: 15 tons (13.6 tonnes) Low P. App.: 8 tons (7.3 tonnes) (including skids)	For comparison only.
Historical Multi-Tube CO_2 Module System	5,600 m^2 (60,300 ft^2)	8 ft x 8 ft x 15 ft (2.4 m x 2.4 m x 4.6 m) H x W x L	NA	Ref: [9] Used in 2010 1st Quarterly Report
Prelimary Cylindrical Vessel Design: for Flue Gas[A]	6,000 m^2 (64,600 ft^2)	25 ft x 5 ft (7.6 m x 1.5 m) L x D	Loaded vessel Weight: 7 tons (6.4 tonnes)	Ref. [10]. For use in analysis.
Possible Compact Design: Module Vessels for Flue Gas	1,000 m^2 (10,800 ft^2)	1 m x 1 m x 1 m (3.3 x 3.3 x 3.3 ft) H x W x L	<2 tons (<1.8 tonnes)	Easily stacked boxes. To be refined in future.

Notes:

A. Data is used for the design basis.

Exhibit 3-4: Required Membrane Area [11]

Case	Module A Area	Module B Area	Module C Area	Notes
MTR	315,000 (3,390,000 ft^2)	300,000 (3,230,000 ft^2)	1,000 m^2 (10,800 ft^2)	2 bara Feed Pressure to Module A (29 psia)

3.3 Technical Maturity

This study is based on technology that is presently technically feasible, but not necessarily available as commercially-offered equipment. Bringing the required equipment to the commercial market could require some development by an OEM. DOE/NETL does not require these costs to be reflected in the cost estimate. This study is based on the position that Non-Recurring Engineering (NRE) costs are not included in the cost estimate of this study.

Where equipment required or assumed for this application is not commercially available, such equipment is identified as such. Equipment in or near this category include the following:

- CO_2 Membrane. MTR provided membrane performance based on what they were achieving at laboratory scale in May 2015. It is believed that this performance will be commercially available for an n^{th} full-scale capture system. [12]
- Vacuum Compressor. Although the vacuum compressor application is beyond the limits of some suppliers due to the size, gas composition, and/or required efficiency, the modeled compressor is based on a commercially-available model from MAN Turbo. The model performance is in agreement with an independent study of rotating equipment for the MTR process conducted by Trimeric. [13]
- Flue Gas Compressors. Although the FG compressor application is beyond the limits of some suppliers due to the size, and/or gas composition, the model is based on a commercially-available model from Dresser Rand.
- Flue Gas Expander. WorleyParsons did not receive confirmation of commercial availability for this high volumetric flow, moderate-temperature, low-pressure expander. However, there is no reason such an expander couldn't be developed for a commercial market. For this present analysis, an efficiency of 87% was utilized.
- Low-Temperature Expander. The low temperature expander is a custom design item, and is common to cryogenic processes such as an air separation unit. WorleyParsons obtained budgetary cost estimates for a single expander. The NRE costs were excluded in the present analysis.
- The CO_2 membrane for a supercritical pulverized coal plant is a novel application. No commercial-scale units are in operation. A small pilot unit is in testing at the National Carbon Capture Center. [14] The cost of the membrane units will be priced to exclude the NRE costs.

3.4 Design Cases

The matrix summarizing the study design cases, and system assumptions is presented in Exhibit 3-5. Further details can be found by examining Case 11 and Case 12 of the BBS. [1]

Exhibit 3-5: Design Cases Summary

Parameter	BBS Case 11	BBS Case 12	BBS Case 11 with MTR Capture
Steam Cycle, MPa / °C / °C (psia / °F / °F)	24.1 / 593 / 593 (3,500/1,100/1,100)	24.1 / 593 / 593 (3,500/1,100/1,100)	24.1 / 593 / 593 (3,500/1,100/1,100)
Steam Quality at LP Turbine Exit, %	75.7^{A}	90.7	91.4
Condenser Pressure, mm Hg (in Hg)	51 (2.0)	51 (2.0)	51 (2.0)
CO_2 Control	N/A	Econamine	MTR
CO_2 Capture, %	N/A	90.2	90.0
Steam Pressure at Capture Extraction, MPa (psia)	N/A (N/A)	0.51 (73.5)	2.14 (310.1)^{B}
SO_2 Polishing Requirement, ppmv	N/A (N/A)	10 (10)	N/A (N/A)

Notes:

A. Based on BBS Case 11. Reported quality from BBS Case 11 of 76% is not in agreement with a converged heat and mass balance. During calibration to Case 11, WorleyParsons corrected for this in part by utilizing an exhaust steam quality of approximately 91%.

B. Extraction steam for TEG dehydration unit reboiler.

3.5 BBS Case 11 Calibration

The Aspen Plus and GateCycle models were calibrated using the H&MB for Case 11 detailed in the BBS report. A comparison of the Case 11 performance summary in the BBS report against the performance summary derived from the WorleyParsons models is presented in Exhibit 3-6.

Exhibit 3-6: Case 11 Performance Comparison

	BBS Case 11	BBS Case 11 (WorleyParsons)
GROSS OUTPUT, kWe		
Steam Turbine	580,400	579,999
CONVENTIONAL AUXILIARY LOADS, kWe		
Coal Handling and Conveying	440	440
Pulverizers	2,780	2,780
Sorbent Handling & Reagent Preparation	890	890
Ash Handling	530	530
Primary Air Fans	1,300	1,249
Forced Draft Fans	1,660	1,778
Induced Draft Fans	7,050	7,028
SCR	50	50

	BBS Case 11	**BBS Case 11 (WorleyParsons)**
Fabric Filter	70	70
Wet FGD	2,970	2,970
Miscellaneous Balance of Plant	2,000	2,000
Steam Turbine Auxiliaries	400	400
Condensate Pumps	800	746
Circulating Water Pump	4,730	4,841
Ground Water Pumps	480	480
Cooling Tower Fans	2,440	2,440
Transformer Losses	1,820	1,819
TOTAL AUXILIARY LOADS, kWe	30,410	30,511
TOTAL NET POWER, kWe	549,990	549,488
Net Plant HHV Efficiency, %	39.3	39.2
Net Plant HHV Heat Rate, kJ/kWh (Btu/kWh)	9,165 (8,687)	9,173 (8,695)
COOLING DUTY, GJ/h (MMBtu/h)		
Steam Turbine Condenser	2,298 (2,178)	2,286 (2,167)
CONSUMABLES		
As-Received Coal, kg/h (lb/h)	185,762 (409,528)	185,762 (409,528)
Thermal Input, kWt (MMBtu/h)	1,400,162 (4,778)	1,400,163 (4,778)
Limestone, kg/h (lb/h)	18,437 (40,646)	18,437 (40,646)
Raw Water Withdrawal, m^3/min (gpm)	20.1 (5,321)	21.1 (5,573)
Raw Water Consumption, m^3/min (gpm)	16.0 (4,227)	15.4 (4,056)

Notes:

A. In the BBS Report Case 11 LP turbine exhaust steam quality is approximately 76%, which is lower than what is typically recommended by ST OEMs (~90%). The LP turbine steam exhaust quality of 91% for the MTR case is within the recommended range and consistent with the BBS Report Case 12.

As illustrated in Exhibit 3-6, WorleyParsons' GateCycle model was calibrated to reproduce BBS Case 11 performance with a sufficient level of accuracy. The same model was utilized as a starting point to simulate the design case of the power plant equipped with MTR CO_2 capture system.

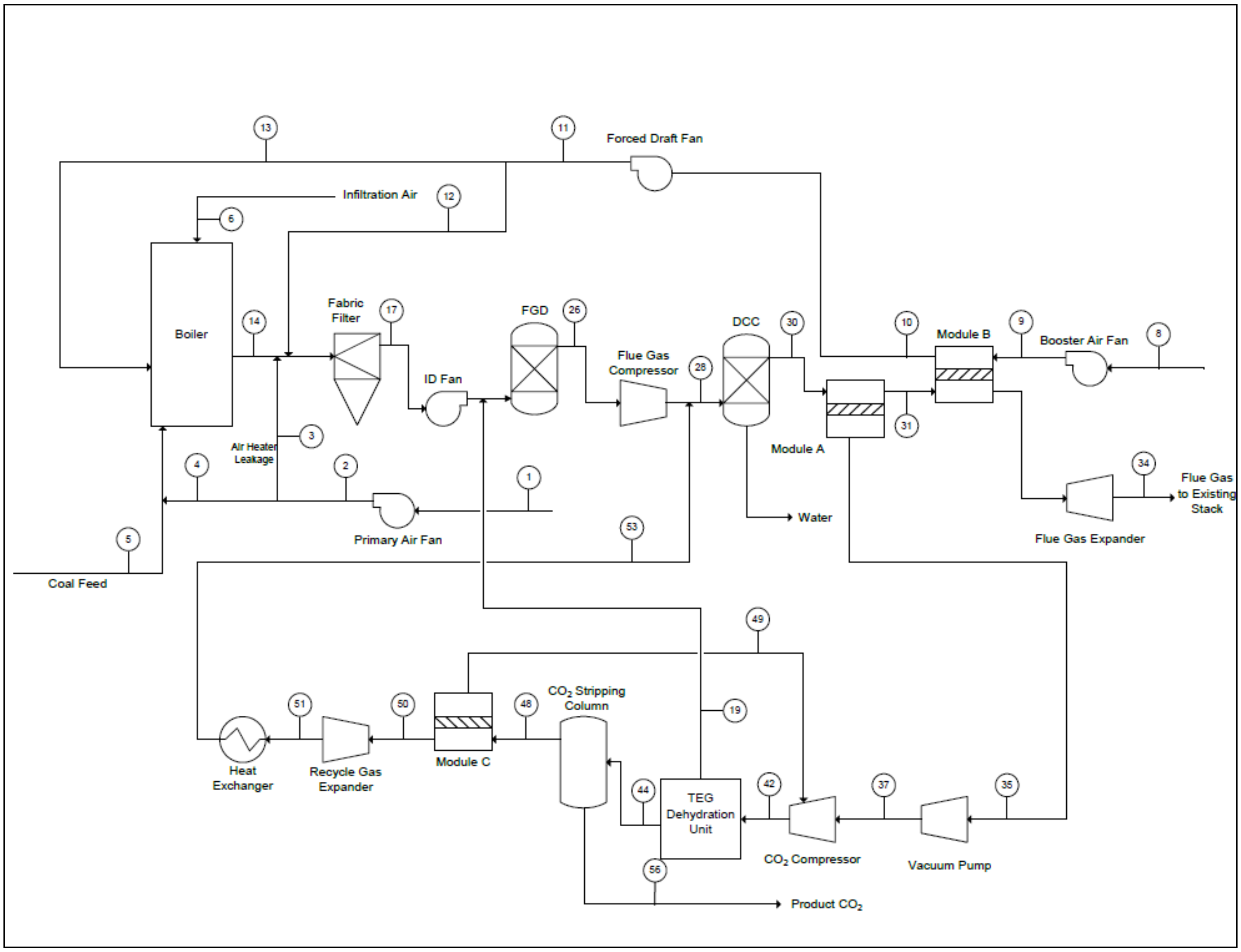
4. SYSTEM DESCRIPTION

The design of the supercritical PC plant equipped with MTR CO_2 capture system is similar to the BBS Case 11, and its description as provided in the BBS report [1]. The MTR CO_2 capture system general process description, process flow diagram, and MTR membrane requirements are provided in Section 3.2 of this report.

Described in this section are those systems that are new to the supercritical PC plant configuration as a result of integrating the MTR CO_2 capture system. Also described in this section are systems that are significantly affected by the CO_2 capture system (particularly the impact of air sweep on the boiler performance and supporting systems).

To aid in this system description is the BFD presented in Exhibit 4-1. Supplementing this section are the heat and mass balances in Appendix 1, the block flow diagram (BFD) and process flow diagrams (PFD) in Appendix 2, Steam Cycle Heat Balance in Appendix 3, and the major equipment list in Section 6.

Exhibit 4-1: Block Flow Diagram of the CO_2 Removal System



4.1 New Systems Required for the CO_2 Membrane Capture System

This section discusses the new systems whose addition is required to support the MTR CO_2 membrane capture system. As compared to the BBS Case 11 supercritical PC plant, the plant equipped with the MTR CO_2 membrane system will require the following additional equipment, systems and modifications.

- Modules A, B, and C of the CO_2 membrane enclosures complete with a gas distribution system and structural supports.
- Flue gas direct contact cooler (DCC) system.
- Flue gas compressor and vacuum compressors.
- Secondary air booster fan.
- Flue gas expanders.
- CO_2 purification and compression system comprised of CO_2 compression/dehydration, chiller and distillation systems.
- Flue gas heat exchangers coupled with a glycol circulating system.
- Process cooling system comprised of additional cooling tower, circulating water system and water treatment capacity to accommodate additional process cooling loads.
- Modifications to the steam cycle system to facilitate a steam extraction for the CO_2 dehydration.

This section provides a description of the new systems added as part of the MTR CO_2 membrane system integration. The corresponding block and process flow diagrams are available in Appendix 2. A major equipment list which characterizes this added equipment is presented in Section 6.

Compared to the BBS Case 11, flue gas system will encounter a recycle stream from the TEG dehydration vent and the new equipment in the ductwork between the ID fans and the FGD absorber. This stream serves to capture CO_2 that would otherwise be lost to venting and to reduce the SO_2 content of the CO_2 product by enabling a small fraction of the CO_2 product to flow through the FGD a second time. Finned heat exchanger tubes have been introduced in the duct work to recover energy from the hot flue gas into a circulating stream of glycol. Energy collected from the flue gas after the ID fan is carried by the glycol to the exit of the counter-flow MTR module, where a second set of finned tubes heat the pressurized flue gas before the flue gas expander. This transfer of energy allows the expander to achieve a greater power output and will maintain additional thermal buoyancy in the flue gas exiting the stack. The lower flue gas temperature entering the FGD absorber also helps to reduce the amount of water which is evaporated.

The desulfurized flue gas leaving the FGD absorber is fed to six compressors operating in parallel. These flue gas compressors raise the pressure up to the 2 bar design pressure of the MTR Module A membranes. The pressurized flue gas is combined with retentate from the MTR crossflow Module C and cooled in a direct contact cooler vessel. In the vessel, which is similar to an FGD absorber, cold water is sprayed over the gas to lower the temperature. Cool, pressurized flue gas is distributed by a header system to the banks of MTR crossflow Module A membranes.

MTR's membranes capture CO_2 by using partial pressure as a driving force across a selective barrier material. The selectivity of the material allows a greater percentage of the CO_2 to preferentially permeate the membrane while those compounds which would be impurities in a CO_2 product preferentially pass through as a retentate stream. A vacuum on the permeate side of these cross-flow membranes provides additional pressure gradient to drive the CO_2 capture.

Retentate from the cross-flow Module A membranes is distributed through banks of MTR counter-flow Module B membranes. In these membranes the CO_2 permeates from the flue gas into the boiler's secondary air. This membrane creates a CO_2 recirculation loop within the plant to ensure that the desired 90% CO_2 capture level is achieved. A booster air fan is added to drive the secondary air through the banks of Module B membranes. The CO_2 depleted flue gas, which exits the Module B membranes, is at a pressure greater than what is required to ensure proper dispersion through the plant stack.

The second set of finned heat exchanger tubes are installed in the ductwork between the crossflow Module B membranes and a flue gas expander. These tubes transfer energy from the recirculating glycol into the flue gas, raising the temperature of the gas. A single stage expander recovers energy from the hot pressurized gas as the pressure is reduced from the operating pressure of the membrane modules to the pressure required to dispel the gas through the stack.

Dry vacuum compressors are utilized to maintain vacuum on the permeate side of the crossflow Module A membranes. Liquid ring vacuum pumps which are used at power plants to maintain the vacuum in the condenser are not well suited to this membrane service. The CO_2 and SO_2 are water soluble gases and water is used in high volumes as a sealant in a liquid ring vacuum pump. Dissolution of CO_2 into the water reduces the systems capture percentage and is undesirable in this application. In addition to the CO_2 losses through the water, auxiliary power consumption by liquid ring pumps is prohibitive to the process. Alternatives to liquid ring vacuum pumps are dry-type compressors which are used to achieve large volumes of vacuum in the pulp and paper industry and are better suited to MTR's process.

Dry compressors operate more efficiently and consume less power. The vacuum compressors for this large volume service would be designed with a combination of axial and radial flow stages to achieve the desired level of vacuum. Three large vacuum compressors are required by the process to capture permeate from MTR crossflow Module A. The CO_2 rich permeate from cross-flow Module A also contains water and SO_2 which have an influence on the vacuum pump materials of construction. At the discharge of the vacuum pump system the CO_2-rich gas is cooled, water is removed, and the gas is piped to machines which will perform further compression.

Two multi-stage compressors operating in parallel will be required to process the volume of gas present following the vacuum compressors. Permeate from MTR's crossflow Module C is introduced into one of the later stages of compression. Circulating water from the cooling tower will be supplied to intercoolers between the compression stages to remove heat which is generated by the compression process. Efficient intercooling reduces the auxiliary power consumption of the machines. Pressurized CO_2-rich gas exiting the compressors must be dried and purified before being pumped to the final boundary limit pressure.

A tri-ethylene glycol (TEG) drying system is installed following the compression system to remove moisture which was not knocked-out in the compression process. The TEG is a temperature swing chemical absorption system in which lean and rich TEG solvent is circulated between the regenerator and absorber. For this application where less than 500 ppmv of H_2O in the product gas is required, a high dew point depression TEG dehydration process has been utilized.

Dry CO_2 rich gas still contains impurities which exceed the values specified in Exhibit 5-2 and must be further treated. Purification of the CO_2-rich gas to produce CO_2 that meets the specified requirements is done utilizing a low-temperature partial condensation process integrated with a distillation column. Cooling water has already been utilized to reduce the temperature and other heat sinks must be used. The gas exiting the CO_2 drying system is cooled down to the necessary temperature in two stages. The relatively hot CO_2-rich gas leaving the drying system is used to meet the energy demands of the CO_2 stripping column reboiler in the first stage of cooling, which eliminates a process steam demand. A refrigeration system based on evaporation of liquid propane is used to reduce the temperature further and partially condense the CO_2 rich gas.

Condensation of the CO_2 dominates at the design temperature and pressure of the gas condenser (propane evaporator). However, oxygen and nitrogen condense with the CO_2 at levels that exceed the product specification. Impurities in the CO_2 are removed by processing the liquid mixture in a stripping column. As the impure liquid CO_2 cascades down the column, vapor which is generated in the column reboiler travels upwards. The impurities preferentially fractionate into the vapor phase as it moves up and out of the column. A pure liquid CO_2 product which meets all specifications[B] is drawn off of the bottom and pumped up to the final discharge pressure.

Overheads from the CO_2 stripping column contain a residual fraction of CO_2. MTR's crossflow Module C recovers a portion of the CO_2 and returns it to the compression system. The pressure differential between the column overheads and the suction pressure of the compressor stage is used as the driving force for permeation. The retentate gas which passes through this crossflow module is at high pressure and still contains CO_2. Power is recovered from the membrane retentate through a low-temperature expander. The expanded retentate is cold and must be warmed prior to reintroduction into the flue gas. After heat exchange with the propane refrigeration system, the warmed retentate is reintroduced into the flue gas prior to the direct contact cooler to give the overall system another chance to capture the recycled CO_2.

A chilling system which utilizes propane as the refrigerant is supplied to achieve the temperatures necessary to condense the CO_2 mixture. Gaseous propane is compressed up to a pressure which will facilitate condensation at a temperature that can be achieved by cooling water. The efficiency of the chilling system is increased by using the liquid propane to reject heat to the purification process. Expanded Module C retentate, product CO_2, and stripping column overheads are all at temperatures lower than that of the liquid propane. Through heat exchange with these three gaseous streams, the liquid propane can be sub-cooled. This process

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[B] TEG dehydration is used extensively for drying natural gas, but has much less maturity for CO_2 applications. A more rigorous analysis and discussion with TEG dehydration system vendors is recommended to ensure the final CO_2 product contains less than the specified 174 ppbv of water.[15]

reduces the losses associated with the refrigeration cycle. At the reduced pressure the liquid-vapor mixture of propane is sent to the propane evaporator where the liquid propane evaporates inducing condensation in the CO_2-rich gas flowing to the purification column.

Additional cooling tower and auxiliary cooling water capacity (compared to a plant with no CO_2 capture) is required to meet the process cooling demands of the MTR CO_2 capture and compression systems. This additional circulating water capacity will service the vacuum compressors, multistage compressors, direct contact cooler, and propane compressor systems. Makeup water demand for the new cooling tower will be offset by collecting the condensate from the DCC and the compression process and pumping it to the cooling tower basin.

4.2 Base BBS Case 11 Systems Impacted by the CO_2 Membrane System

This section discusses those systems that are impacted in a substantial manner as a result of the integration of MTR CO_2 membrane capture system.

Addition of the MTR CO_2 capture system to the BBS Case 11 supercritical PC plant will reduce plant efficiency, thus, requiring an increase in plant gross power generation to maintain the same nominal net power output of 550 MW as in BBS Case 11. As a result, the capacities of all of the plant systems need to be increased, including:

- Steam generator and steam turbine systems,
- Feed water and condensate systems,
- Cooling system,
- Material handling systems,
- Water and waste water treatment,
- High, medium, and low voltage electrical systems,
- Environmental controls systems,
- Piping systems,
- Instrumentation and control system,
- Larger structures and buildings and
- Larger (all other) plant auxiliary systems.

In addition, the incorporation of the air sweep recycle will impact the design of the steam generator and downstream components. These changes will be discussed in the subSections below.

4.2.1 Steam Generator

The steam generator secondary air system is used as a sweep gas in the MTR counterflow Module B to enhance CO_2 capture. This results in the vitiated air from Module B being fed into the secondary air system and an increased mass and volumetric flow rate, as presented in Exhibit 4-2.

Exhibit 4-2: Changes in Steam Generator and Secondary Air System

Parameter	BBS Case 11	BBS Case 11 with MTR Capture
Secondary Air to Windbox		
Mass Flow Rate, kg/h (lb/h)	1,355,712 (2,988,782)	2,187,201 (4,821,872)
Volumetric Flow Rate, Nm^3/min (MMscfd)	18,612 (946)	28,842 (1,467)
CO_2 Content, mol%	0.03	7.56
O_2 Content, mol%	20.74	18.40
Molecular Weight	28.86	29.94
Boiler Excess Air, %[C]	8.19	8.28
Boiler Efficiency, %	88.00	86.23

In December of 2013, B&W issued a report to MTR reviewing the effect of CO_2 enriched air on the combustion performance of a steam generator in order to simulate the integration of the MTR CO_2 membranes into the PC power plant. While that report is not yet public, the key results of that report have been incorporated into this evaluation in order to develop realistic simulation results. Key assumptions from that report that have been incorporated herein are listed below:

- A reduced windbox O_2 concentration of approximately 18% as suggested by B&W to be a near optimum operating point.
- A boiler fuel efficiency degradation of about 2%.

In addition to these assumptions WorleyParsons also assumed the following:

- The primary air to coal flow ratio was unchanged due to its essential role in proper coal transport, drying and combustion.
- The volumetric flow of the secondary air was allowed to increase to compensate for the reduced O_2 level. The flow of the secondary air was set to preserve the excess air utilized in BBS Case 11.
- Proper design and tuning will allow the NO_X and CO emissions to remain essentially unchanged from the BBS Case 11 values. Without redesign, the addition of the CO_2 diluent would reduce the furnace temperature and lower NOx while increasing CO emissions. Redesigning the boiler should allow removal of some “Low NO_X features”, thus allowing the NO_X level to rise and the CO level to drop.

[C] The boiler excess air is typically 15 to 20% as opposed to the 8% found in the BBS Case 11. Since the MTR application will be compared to the BBS Case 11, this analysis has continued with a similar 8% excess analysis.

4.2.2 Particulate Removal System

The PM collection efficiency of the fabric filter is assumed to be consistent with that of BBS Case 11 or 99.8%. The PM leaving the plant stack is expected to be lower than that of BBS Case 11 as PM leaving the fabric filter would pass through additional mitigating equipment. Specifically, downstream of the fabric filter is the FGD and DCC, which are expected to reduce the PM levels by approximately 75% and 40 to 50%, respectively. Downstream of the DCC are the membrane modules. From a plant emission point of view, the emitted PM will be lower than BBS Case 11 because of the newly added DCC and membrane system.

This anticipated performance change for the particulate removal system is not reported in the performance table as the BBS did not reflect any PM downstream of the fabric filter.

4.2.3 Steam Cycle

Electrical energy is the prime driver for the equipment which would be installed to capture CO_2 with MTR's membranes. As such, only a minor modification to the steam cycle is required as a result of the new systems. Intermediate pressure steam is used to provide the heat required to dry the CO_2 before it is liquefied. A pipe would divert IP steam to the drying unit from the IP turbine extraction piping serving feedwater heater number 6. This steam will be desuperheated with boiler feedwater. Steam condensate from the dryer is returned to the plant's deaerator.

5. TECHNICAL ANALYSIS RESULTS

The plant configuration for the MTR Case is the same as BBS Case 11 with the exception that the MTR CO_2 membrane separation, purification and compression systems have been added for CO_2 capture. The nominal net output is maintained at 550 MW by increasing the boiler size and turbine/generator size to account for the greater auxiliary load imposed by the CO_2 capture systems.

The configuration block flow diagram (BFD) and process flow diagram (PFD) of the MTR process, and the flue gas side of the base power plant are presented in Appendix 1. The PFD includes the CO_2 membrane system, CO_2 compression and purification, and all major fans, blowers, vacuum compressors, cooling, etc.

The heat and mass balance (H&MB) stream tables for the MTR Case are found in Appendix 2. This table covers the boiler, and CO_2 membrane system, and all systems except for the steam turbine cycle which is presented in Appendix 3.

A performance comparison of the MTR design case to the BBS Cases 11 and 12 is presented in Exhibit 5-1.

Exhibit 5-1: Comparative Plant Performance Summary

	BBS Case 11	BBS Case 12	BBS Case 11 with MTR Capture
GROSS OUTPUT, kWe			
Steam Turbine	580,400	662,800	780,795
CONVENTIONAL AUXILIARY LOADS, kWe			
Coal Handling and Conveying	440	510	608
Pulverizers	2,780	3,850	3,842
Sorbent Handling & Reagent Preparation	890	1,250	1,230
Ash Handling	530	740	732
Primary Air Fans	1,300	1,800	1,726
Forced Draft Fans	1,660	2,300	334
Induced Draft Fans	7,050	11,120	11,594
SCR	50	70	69
Fabric Filter	70	100	97
Wet FGD	2,970	4,110	4,104
Miscellaneous Balance of Plant[B,C]	2,000	2,000	2,000
Steam Turbine Auxiliaries	400	400	400
Condensate Pumps	800	560	1,005
Circulating Water Pump	4,730	10,100	4,955
Ground Water Pumps	480	910	663
Cooling Tower Fans	2,440	5,230	4,873
Transformer Losses	1,820	2,290	2,563

CARBON CAPTURE AUXILIARY LOADS, kWe			
Flue Gas Compressor	0	0	77,869
Direct Contact Cooler	0	0	342
Flue Gas Expander	0	0	-33,279
Module C Recycle Gas Expander	0	0	-3,337
Secondary Air Booster Fan	0	0	5,238
Vacuum Compressor^D	0	20,600	37,475
CO_2 Compressor	0	44,890	68,842
CO_2 Dehydration Unit	0	0	136
Propane Refrigeration Compressor	0	0	24,672
CO_2 Pump	0	0	3,795
Auxiliary Cooling Pumps	0	0	4,756
TOTAL AUXILIARY LOADS, kWe	30,410	112,830	227,303
TOTAL NET POWER, kWe	549,990	549,970	553,492
Net Plant HHV Efficiency, %	39.3	28.4	28.6
Net Plant HHV Heat Rate, kJ/kWh (Btu/kWh)	9,165 (8,687)	12,663 (12,002)	12,585 (11,929)
COOLING DUTY, GJ/h (MMBtu/h)			
Steam Turbine Condenser	2,298 (2,178)	1,737 (1,646)	3,077 (2,916)
Auxiliary Cooling Users	0 (0)	0 (0)	1,489 (1,411)
CONSUMABLES			
As-Received Coal, kg/h (lb/h)	185,762 (409,528)	256,656 (565,820)	256,715 (565,950)
Thermal Input^A, kWt (MMBtu/h)	1,400,162 (4,778)	1,934,519 (6,601)	1,934,964 (6,602)
Limestone, kg/h (lb/h)	18,437 (40,646)	25,966 (57,245)	26,647 (58,745)
Raw Water Withdrawal, m^3/min (gpm)	20.1 (5,321)	38.1 (10,071)	33 (8,744)
Raw Water Consumption, m^3/min (gpm)	16.0 (4,227)	29.3 (7,733)	24 (6,302)

Notes:

A. HHV of as-received Illinois No. 6 coal is 27,135 kJ/kg (11,666 Btu/lb)
B. Boiler feed pumps are turbine driven
C. Includes plant control systems, lighting, HVAC and miscellaneous low-voltage loads
D. Represents Econamine auxiliary loads for BBS Case 12
E. Pressure and temperature loss for steam from steam turbine to TEG dehydration unit is not accounted for, similar to the BBS report.
F. BBS Case 11 has an LP turbine exhaust steam quality of approximately 76%, which is lower than what is typically recommended by steam turbine OEM's (~90%). The LP turbine steam exhaust quality of approximately 91% for MTR case is within the recommended range and consistent with BBS Case 12.

The CO_2 pipeline specification [1], [15], and estimated characteristics of the product CO_2 at the pipeline inlet are presented in Exhibit 5-2.

Exhibit 5-2: Product CO_2 Characteristics

Parameter	NETL Quality Guideline	BBS Case 11 with MTR Capture
Pressure, MPa (psia)	15.27 (2,215)	15.27 (2,215)
Maximum Ar, vol%	1.0	0.0
Mininum CO_2, vol%	95.00	99.86
Maximum N_2, vol%	1.0	0.0
Maximum H_2O, ppmv	500	449
Maximum O_2, ppmv	10.0	9.9
Maximum SO_2, ppmv	100.0	100.0

Flue gas analysis and emissions are presented in Exhibit 5-3

Exhibit 5-3: Flue Gas Analysis and Emissions

	kg/GJ (lb/MMBtu)	tonne/y (ton/y)[A]	kg/MWh (lb/MWh) Gross	kg/MWh (lb/MWh) Net
SO_2	0.001 (0.002)	48 (53)	0.008 (0.018)	0.012 (0.026)
NO_X[B]	0.027 (0.062)	1,383 (1,524)	0.238 (0.524)	0.338 (0.744)
Particulates[3]	0.006 (0.013)	297 (327)	0.051 (0.112)	0.072 (0.160)
Hg	4.91E-7 (1.14E-6)	0.025 (0.028)	4.38E-6 (9.66E-6)	6.22E-6 (1.37E-5)
CO_2	9.0 (20.9)	465,947 (513,610)	80 (177)	114 (251)

Notes:

A. Assumes 85% capacity factor.
B. The mass basis for NO_X is NO_2.
C. Particulate removal is based on 99.8% removal in the fabric filter. In order to agree with BBS reporting, the heat and mass balance tables in Appendix 1 report 100% removal. Particulate removal does not take credit for the additional removal effects of the FGD, direct contact cooler, and membrane modules.

Carbon, sulfur, water and overall energy balances are presented in Exhibit 5-4, Exhibit 5-5, Exhibit 5-6 and Exhibit 5-7.

Exhibit 5-4: Carbon Balance

Carbon In, kg/h (lb/h)		Carbon Out, kg/h (lb/h)	
Coal	163,656 (360,793)	Stack Gas	17,078 (37,650)
Primary Air	81 (178)	FGD Product	641 (1,412)
Infiltration Air	7 (15)	CO_2 Product	149,330 (329,210)
Secondary Air	273 (603)	Direct Contact Cooler Water	1 (2)
FGD Reagent	3,038 (6,697)	Vacuum Compressor Water	0 (0)
FGD Oxidation Air	4 (9)	CO_2 Compressor Water	1 (3)
		Miscellaneous Small Losses	8 (18)
Total	167,059 (368,295)		167,059 (368,295)

Exhibit 5-5: Sulfur Balance

Sulfur In, kg/h (lb/h)		Sulfur Out, kg/h (lb/h)	
Coal	6,444 (14,205)	Stack Gas	3 (7)
		FGD Water	0 (0)
		FGD Product	6,400 (14,109)
		CO_2 Product	40 (88)
		Miscellaneous Small Losses	0 (1)
Total	6,444 (14,205)		6,443 (14,205)

Exhibit 5-6: Water Balance

Water Use	Water Demand, m^3/min (gpm)	Internal Recycle, m^3/min (gpm)	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
DCC	0 (0)	4 (1,145)	-4 (-1,145)	0 (0)	-4 (-1,145)
CO_2 Compression and Drying	0 (0)	1 (143)	-1 (-143)	0 (0)	-1 (-143)
FGD Make-Up	5 (1,373)	3 (719)	2 (654)	0 (0)	2 (654)
BFW Make-Up	0 (0)	0 (0)	0 (0)	0 (0)	0 (0)
Cooling Tower	35 (9,273)	0 (23)	35 (9,249)	9 (2,314)	27 (6,936)
Miscellaneous Users	1 (238)	0 (110)	0 (129)	0 (129)	0 (0)
Total	41 (10,884)	8 (2,140)	33 (8,744)	9 (2,443)	24 (6,302)

Exhibit 5-7: Overall Energy Balance

	HHV	Sensible and Latent[A]	Power	Total
Heat In, GJ/h (MMBtu/h)				
Coal	6,966 (6,602)	-540 (-512)	0 (0)	6,426 (6,090)
Air	0 (0)	-261 (-247)	0 (0)	-261 (-247)
Raw Water Make-Up	0 (0)	-36,862 (-34,939)	0 (0)	-36,862 (-34,939)
CO_2 Capture System Blowdown to Cooling Tower	0 (0)	-4,680 (-4,436)	0 (0)	-4,680 (-4,436)
FGD Reagent [B]	0 (0)	-320 (-304)	0 (0)	-320 (-304)
Auxiliary Power		0 (0)	818 (776)	818 (776)
Totals	6,966 (6,602)	-42,664 (-40,438)	818 (776)	-34,880 (-33,060)
Heat Out, GJ/h (MMBtu/h)				
Bottom Ash	0 (0)	-2 (-2)	0 (0)	-2 (-2)
Fly Ash	0 (0)	-13 (-12)	0 (0)	-13 (-12)
FGD Product	0 (0)	-482 (-456)	0 (0)	-482 (-456)
Flue Gas	0 (0)	-603 (-571)	0 (0)	-603 (-571)
Cooling Tower[C]	0 (0)	-20,731 (-19,649)	0 (0)	-20,731 (-19,649)
CO_2 Product	0 (0)	-5,043 (-4,780)	0 (0)	-5,043 (-4,780)
Cooling Tower Blowdown to Discharge	0 (0)	-8,411 (-7,972)	0 (0)	-8,411 (-7,972)
FGD Blowdown to Discharge	0 (0)	-2,376 (-2,252)	0 (0)	-2,376 (-2,252)
Sanitary Water To Discharge	0 (0)	-18 (-17)	0 (0)	-18 (-17)
Ash Sluice Water to Discharge	0 (0)	-20 (-19)	0 (0)	-20 (-19)
Steam and Sampling Water Losses	0 (0)	-212 (-201)	0 (0)	-212 (-201)
Service Water Users to Discharge	0 (0)	-218 (-207)	0 (0)	-218 (-207)
Boiler Heat Losses	0 (0)	80 (76)	0 (0)	80 (76)
Process Losses	0 (0)	359 (340)	0 (0)	359 (340)
Power	0 (0)	0 (0)	2,811 (2,664)	2,811 (2,664)
Totals	0 (0)	-37,691 (-35,724)	2,811 (2,664)	-34,880 (-33,060)

Notes:

A. Stream enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.
B. Dry reagent. Slurry mixing water is accounted for in the raw water make-up.
C. Includes all cooling water, both form the power plant and the CO_2 capture system.

6. MAJOR EQUIPMENT LIST

The resulting mass and energy balance data from the simulation models were used to size major pieces of equipment.

Major equipment lists for the supercritical PC plant equipped with MTR CO_2 capture system is shown in the following tables. The accounts scope and numbers, and design assumptions used in this equipment list are consistent with the BBS report Case 11.

Account 1 - Fuel and Sorbent Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	181 tonne (200 ton)	2	0
2	Feeder	Belt	572 tonne/h (630 tph)	2	0
3	Conveyor No. 1	Belt	1,134 tonne/h (1,250 tph)	1	0
4	Transfer Tower No. 1	Enclosed	N/A	1	0
5	Conveyor No. 2	Belt	1,134 tonne/h (1,250 tph)	1	0
6	As-Received Coal Sampling System	Two-stage	N/A	1	0
7	Stacker/Reclaimer	Traveling, linear	1,134 tonne/h (1,250 tph)	1	0
8	Reclaim Hopper	N/A	54 tonne (60 ton)	2	1
9	Feeder	Vibratory	209 tonne/h (230 tph)	2	1
10	Conveyor No. 3	Belt w/ tripper	426 tonne/h (470 tph)	1	0
11	Crusher Tower	N/A	N/A	1	0
12	Coal Surge Bin w/ Vent Filter	Dual outlet	209 tonne (230 ton)	2	0
13	Crusher	Impactor reduction	8 cm x 0 - 3 cm x 0 (3" x 0 - 1-1/4" x 0)	2	0
14	As-Fired Coal Sampling System	Swing hammer	N/A	1	1
15	Conveyor No. 4	Belt w/tripper	426 tonne/h (470 tph)	1	0
16	Transfer Tower No. 2	Enclosed	N/A	1	0
17	Conveyor No. 5	Belt w/ tripper	426 tonne/h (470 tph)	1	0
18	Coal Silo w/ Vent Filter and Slide Gates	Field erected	907 tonne (1,000 ton)	3	0
19	Limestone Truck Unloading Hopper	N/A	36 tonne (40 ton)	1	0
20	Limestone Feeder	Belt	109 tonne/h (120 tph)	1	0
21	Limestone	Belt	109 tonne/h (120 tph)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
	Conveyor No. L1				
22	Limestone Reclaim Hopper	N/A	18 tonne (20 ton)	1	0
23	Limestone Reclaim Feeder	Belt	91 tonne/h (100 tph)	1	0
24	Limestone Conveyor No. L2	Belt	91 tonne/h (100 tph)	1	0
25	Limestone Day Bin	w/ actuator	354 tonne (390 ton)	2	0

Account 2 – Coal and Sorbent Preparation and Feed

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Coal Feeder	Gravimetric	45 tonne/h (50 tph)	6	0
2	Coal Pulverizer	Ball type or equivalent	45 tonne/h (50 tph)	6	0
3	Limestone Weigh Feeder	Gravimetric	29 tonne/h (32 tph)	1	1
4	Limestone Ball Mill	Rotary	29 tonne/h (32 tph)	1	1
5	Limestone Mill Slurry Tank with Agitator	N/A	113,563 liters (30,000 gal)	1	1
6	Limestone Mill Recycle Pumps	Horizontal centrifugal	1,893 lpm @ 12 m H2O (500 gpm @ 40 ft H2O)	1	1
7	Hydroclone Classifier	4 active cyclones in a 5 cyclone bank	492 lpm (130 gpm) per cyclone	1	1
8	Distribution Box	2-way	N/A	1	1
9	Limestone Slurry Storage Tank with Agitator	Field erected	632,169 liters (167,000 gal)	1	1
10	Limestone Slurry Feed Pumps	Horizontal centrifugal	1,325 lpm @ 9 m H2O (350 gpm @ 30 ft H2O)	1	1

Account 3 – Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1,446,039 liters (382,000 gal)	2	0
2	Condensate Pumps	Vertical canned	31,041 lpm @ 213 m H2O (8,200 gpm @ 700 ft H2O)	1	1
3	Deaerator and Storage Tank	Horizontal spray type	2,410,846 kg/h (5,315,000 lb/h), 5 min. tank	1	0
4	Boiler Feed Pump/Turbine	Barrel type, multi-stage, centrifugal	40,504 lpm @ 3,475 m H2O (10,700 gpm @ 11,400 ft H2O)	1	1
5	Startup Boiler Feed Pump, Electric Motor Driven	Barrel type, multi-stage, centrifugal	12,113 lpm @ 3,475 m H2O (3,200 gpm @ 11,400 ft H2O)	1	0
6	LP Feedwater Heater 1A/1B	Horizontal U-tube	920,794 kg/h (2,030,000 lb/h)	2	0
7	LP Feedwater Heater 2A/2B	Horizontal U-tube	920,794 kg/h (2,030,000 lb/h)	2	0
8	LP Feedwater Heater 3A/3B	Horizontal U-tube	920,794 kg/h (2,030,000 lb/h)	2	0
9	LP Feedwater Heater 4A/4B	Horizontal U-tube	920,794 kg/h (2,030,000 lb/h)	2	0
10	HP Feedwater Heater 6	Horizontal U-tube	2,408,578 kg/h (5,310,000 lb/h)	1	0
11	HP Feedwater Heater 7	Horizontal U-tube	2,408,578 kg/h (5,310,000 lb/h)	1	0
12	HP Feedwater heater 8	Horizontal U-tube	2,408,578 kg/h (5,310,000 lb/h)	1	0
13	Auxiliary Boiler	Shop fabricated, water tube	18,144 kg/h, 2.8 MPa, 343 °C (40,000 lb/h, 400 psig, 650 °F)	1	0
14	Fuel Oil System	No. 2 fuel oil for light off	1,135,632 liter (300,000 gal)	1	0
15	Service Air Compressors	Flooded Screw	28 m3/min @ 0.7 MPa (1,000 scfm @ 100 psig)	2	1
16	Instrument Air Dryers	Duplex, regenerative	28 m3/min (1,000 scfm)	2	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
17	Closed Cycle Cooling Heat Exchangers	Shell and tube	53 MMkJ/h (50 MMBtu/h) each	2	0
18	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	293,372 lpm @ 30 m H2O (77,500 gpm @ 100 ft H2O)	2	1
19	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3,785 lpm @ 88 m H2O (1,000 gpm @ 290 ft H2O)	1	1
20	Fire Service Booster Pump	Two-stage horizontal centrifugal	2,650 lpm @ 64 m H2O (700 gpm @ 210 ft H2O)	1	1
21	Raw Water Pumps	Stainless steel, single suction	9,501 lpm @ 18 m H2O (2,510 gpm @ 60 ft H2O)	2	1
22	Ground Water Pumps	Stainless steel, single suction	3,785 lpm @ 268 m H2O (1,000 gpm @ 880 ft H2O)	5	1
23	Filtered Water Pumps	Stainless steel, single suction	2,688 lpm @ 49 m H2O (710 gpm @ 160 ft H2O)	2	1
24	Filtered Water Tank	Vertical, cylindrical	2,577,885 liter (681,000 gal)	1	0
25	Makeup Water Demineralizer	Multi-media filter, cartridge filter, RO membrane assembly, electrodeionization unit	795 lpm (210 gpm)	1	1
26	Liquid Waste Treatment System	--	10 years, 24-hour storm	1	0

Account 4 – Boiler and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Boiler	Supercritical, drum, wall-fired, low NOx burners, overfire air	2,408,578 kg/h steam @ 25.5 MPa/602 °C/602 °C (5,310,000 lb/h steam @ 3,700 psig/1,115 °F/1,115 °F)	1	0
2	Primary Air Fan	Centrifugal	326,133 kg/h, 4,451 m3/min @ 119 cm WG (719,000 lb/h, 157,200 acfm @ 47 in. WG)	2	0
3	Forced Draft Fan	Centrifugal	1,239,669 kg/h, 16,795 m3/min @ 6 cm WG (2,733,000 lb/h, 593,100 acfm @ 2 in. WG)	2	0
4	Induced Draft Fan	Centrifugal	1,721,385 kg/h, 38,715 m3/min @ 94 cm WG (3,795,000 lb/h, 1,367,200 acfm @ 37 in. WG)	2	0
5	SCR Reactor Vessel	Space for spare layer	3,302,156 kg/h (7,280,000 lb/h)	2	0
6	SCR Catalyst	--	--	3	0
7	Dilution Air Blower	Centrifugal	119 m3/min @ 108 cm WG (4,200 acfm @ 42 in. WG)	2	1
8	Ammonia Storage	Horizontal tank	132,490 liter (35,000 gal)	5	0
9	Ammonia Feed Pump	Centrifugal	25 lpm @ 91 m H2O (7 gpm @ 300 ft H2O)	2	1

Account 5 – Flue Gas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Fabric Filter	Single stage, high-ratio with pulse-jet online cleaning system	1,732,271 kg/h (3,819,000 lb/h) 99.8% efficiency	2	0
2	FGD Absorber Module	Counter-current open spray	55,558 m3/min (1,962,000 acfm)	1	0
3	Recirculation Pumps	Horizontal centrifugal	193,058 lpm @ 64 m H2O (51,000 gpm @ 210 ft H2O)	5	1
4	Bleed Pumps	Horizontal centrifugal	5,716 lpm (1,510 gpm) at 20 wt% solids	2	1
5	Oxidation Air Blowers	Centrifugal	219 m3/min @ 0.3 MPa (7,730 acfm @ 37 psia)	2	1
6	Agitators	Side entering	50 hp	5	1
7	Dewatering Cyclones	Radial assembly, 5 units each	1,438 lpm (380 gpm) per cyclone	2	0
8	Vacuum Filter Belt	Horizontal belt	45 tonne/h (50 tph) of 50 wt % slurry	2	1
9	Filtrate Water Return Pumps	Horizontal centrifugal	871 lpm @ 12 m H2O (230 gpm @ 40 ft H2O)	1	1
10	Filtrate Water Return Storage Tank	Vertical, lined	567,816 lpm (150,000 gal)	1	0
11	Process Makeup Water Pumps	Horizontal centrifugal	4,580 lpm @ 21 m H2O (1,210 gpm @ 70 ft H2O)	1	1

Account 5B – CO_2 Removal and Compression

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Flue Gas Glycol Cooler	Finned tube	188 GJ/h (179 MMBtu/h), 11,768 m2 (126,673 ft2), Gas side 0.11 MPa (15.3 psia) / 223 °C (433 °F) / SS, Tube side 0.40 MPa (58.0 psia) / 154 °C (310 °F) / SS	1	0
2	Flue Gas Compressor	Centrifugal	8,417 m3/min (297,257 acfm), Inlet pressure 0.10 MPa (14.7 psia), Outlet pressure 0.20 MPa (29.4 psia), 12,978 kW (17,404 hp), Material CS, Impeller material SS	6	0
3	Direct Contact Cooler	N/A	32,343 m3/min (1,142,166 acfm), 1009 GJ/h (956 MMBtu/h), 0.20 MPa (29.0 psia) /, 130 °C (266°F)	1	0
4	Flue Gas Glycol Heater	Finned tube	188 GJ/h (179 MMBtu/h), 10,440 m2 (112,373 ft2), Gas side 0.19 MPa (27.6 psia) / 107 °C (225 °F) / SS, Tube side 0.40 MPa (58.0 psia) / 154 °C (310 °F) / SS	1	0
5	Glycol Circulation Pump	N/A	9,431 lpm @ 30 m H2O, (2,491 gpm @ 100 ft H2O), Material CS, Impeller material SS	3	1
6	Flue Gas Expander	Centrifugal	21,549 m3/min (760,990 acfm), Inlet pressure 0.19 MPa (27.5 psia), Outlet pressure 0.10 MPa (14.7 psia), 33,279 kW (44,628 hp), Material CS, Impeller material SS	1	0
7	Booster Air Fan	Centrifugal	27,321 m3/min (964,815 acfm), 0.10 MPa (14.7 psia), 0.11 MPa (16.0 psia), 5,238 kW (7,025 hp), Material CS, Impeller material CS	1	0
8	MTR CO2 Module A	Cross flow Polaris membrane	21922 m3/min (774,174 acfm), CO2 feed concentration 20.4 mol% 315,000 m2 (3,390,632 ft2)	1	0
9	MTR CO2 Module B	Counter flow Polaris	18232 m3/min (643,873 acfm), CO2 feed concentration 8.5 mol% 300,000 m2 (3,229,173 ft2)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
		membrane			
10	MTR CO2 Module C	Cross flow Polaris membrane	131 m3/min (4,638 acfm), CO2 feed concentration 58.1 mol% 1,000 m2 (10,764 ft2)	1	0
11	Vacuum compressor with Inter-Stage Cooling	Axial-radial	13,911 m3/min (491,280 acfm), Inlet pressure 0.02 MPa (2.9 psia), Outlet pressure 0.12 MPa (16.9 psia), 12,492 kW (16,751 hp), Material CS, Impeller material SS	3	0
12	Raw CO2 Cooler	Shell and tube	34 GJ/h (33 MMBtu/h), 4,533 m2 (48,788 ft2), Shell side 0.12 MPa (16.9 psia) / 66 °C (150 °F) / SS, Tube side 0.45 MPa (64.7 psia) / 27 °C (80 °F) / CS	1	0
13	Raw CO2 Knock Out Drum	Vertical	Diameter 3 m (11 ft), Height (T/T) 4 m (13 ft), 0.11 MPa (16.2 psia), 35 °C (95 °F), SS	3	0
14	Multi-Stage CO2 Compressor with Inter-Stage Cooling	Multi-stage integrally-geared centrifugal	367,762 kg/h (810,763 lb/h), 3,513 m3/min (124,052 acfm), Inlet pressure 0.11 MPa (16.2 psia), Outlet pressure 3.02 MPa (438.6 psia), 34,421 kW (46,159 hp), Material CS, Impeller material SS	2	0
15	CO2 Condensate Pump	Horizontal centrifugal	540 lpm @ 30 m H2O, (143 gpm @ 100 ft H2O), Material CS, Impeller material SS	1	1
16	DCC Blowdown Pump	Horizontal centrifugal	4,333 lpm @ 30 m H2O, (1,145 gpm @ 100 ft H2O), Material CS, Impeller material SS	1	1
17	Module C Recycle Gas Expander	Centrifugal	69 m3/min (2,428 acfm), Inlet pressure 2.67 MPa (387.3 psia), Outlet pressure 0.22 MPa (31.4 psia), 3,337 kW (4,475 hp), Material SS, Impeller material SS	1	0
18	Liquid CO2	Horizon	9,483 lpm (2,505 gpm),	1	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
	Pump	tal centrifugal	548,110 kg/h (1,208,354 lb/h), Inlet pressure 2.91 MPa (422.1 psia), Outlet pressure 15.30 MPa (2219.7 psia), 3,795 kW (5,089 hp), Material SS, Impeller material SS		
19	CO2 Dehydration Package	TEG dehydrator	TEG circulation rate 82,122 kg/h (181,044 lb/h), Gas inlet flow 295 m3/min (10,433 acfm), Gas inlet pressure 3.02 MPa (438.6 psia), Gas inlet water 3,565 ppmv, Gas exit water 242 ppmv	1	0
20	CO2 Stripping Column Reboiler	Shell and tube	41 GJ/h (39 MMBtu/h), 954 m2 (10,264 ft2), Shell side 2.91 MPa (422.1 psia) / -6 °C (21 °F) / SS, Tube side 2.95 MPa (428.4 psia) / 60 °C (141 °F) / SS	1	0
21	Propane Evaporator	Shell and tube	218 GJ/h (207 MMBtu/h), 5429 m2 (58,440 ft2), Shell side 0.14 MPa (20.0 psia) / -35 °C (-31 °F) / SS, Tube side 2.95 MPa (428.3 psia) / 20 °C (67 °F) / SS	1	0
22	CO2 Stripping Column	Diameter	Diameter 4 m (12 ft), Height (T/T) 15 m (51 ft), Actual trays 22 Operating pressure 2.90 MPa (421.0 psia), Operating temperature -29 °C (-20 °F) SS	1	0
23	Propane Chiller Compressor	Centrifugal	3,878 m3/min (136,950 acfm), Inlet pressure 0.14 MPa (19.9 psia), Outlet pressure 1.28 MPa (185.0 psia), 24,672 kW (33,086 hp), Material SS, Impeller material SS	1	0
24	Refrigeration Recovery Exchanger RHX-1	Shell and tube	9 GJ/h (8 MMBtu/h), 468 m2 (5,041 ft2), Shell side 1.28 MPa (185.0 psia) / 53 °C (128 °F) / CS, Tube side 0.21 MPa (30.4 psia) / -15 °C (5 °F) / SS	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
25	Propane Condenser	Shell and tube	254 GJ/h (241 MMBtu/h), 9828 m2 (105,791 ft2), Shell side 1.27 MPa (183.8 psia) / 47 °C (117 °F) / CS, Tube side 0.45 MPa (64.7 psia) / 27 °C (80 °F) / CS	1	0
26	Refrigeration Recovery Exchanger RHX-2	Shell and tube	16 GJ/h (15 MMBtu/h), 1564 m2 (16,834 ft2), Shell side 1.22 MPa (176.8 psia) / 35 °C (95 °F) / CS, Tube side 15.30 MPa (2219.7 psia) / 7 °C (45 °F) / SS	1	0
27	Refrigeration Recovery Exchanger RHX-3	Shell and tube	14 GJ/h (13 MMBtu/h), 636 m2 (6,842 ft2), Shell side 1.21 MPa (175.5 psia) / 28 °C (82 °F) / CS, Tube side 2.90 MPa (421.0 psia) / -29 °C (-20 °F) / SS	1	0
28	Refrigeration Recovery Exchanger RHX-4	Shell and tube	12 GJ/h (11 MMBtu/h), 302 m2 (3,249 ft2), Shell side 1.20 MPa (174.3 psia) / 21 °C (70 °F) / CS, Tube side 0.22 MPa (31.4 psia) / -91 °C (-131 °F) / SS	1	0

Account 6 – Combustion Turbine and Accessories

N/A

Account 7 – HRSG, Ducting & Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	Reinforced concrete with FRP liner	152 m (500 ft) high x 5.8 m (19 ft) diameter	1	0

Account 8 – Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine	Commercially available advanced steam turbine	820 MW, 24.1 MPa/593 °C/593 °C (3500 psig/ 1100 °F/1100 °F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	910 MVA @ 0.9 p.f.,24 kV, 60 Hz	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	3,384 GJ/h (3,210 MMBtu/h), Inlet water temperature 16 °C (60 °F), Water temperature rise 11 °C (20 °F)	1	0

Account 9 – Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	900,935 lpm @ 21.3 m (238,000 gpm @ 70 ft)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi-cell	11 °C (51.5 °F) wet bulb / 16 °C (60 °F) CWT / 27 °C (80 °F) HWT 5,018 GJ/h (4,760 MMBtu/h) heat load	1	0

Account 10 – Ash/Spent Sorbent Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Economizer Hopper (part of boiler scope of supply)	--	--	4	0
2	Bottom Ash Hopper (part of boiler scope of supply)	--	--	2	0
3	Clinker Grinder	--	5.4 tonne/h (6 tph)	1	1
4	Pyrites Hopper (part of pulverizer scope of supply included with boiler)	--	--	6	0
5	Hydroejectors	--	--	12	
6	Economizer /Pyrites Transfer Tank	--	--	1	0
7	Ash Sluice Pumps	Vertical, wet pit	227 lpm @ 17 m H2O (60 gpm @ 56 ft H2O)	1	1
8	Ash Seal Water Pumps	Vertical, wet pit	7,571 lpm @ 9 m H2O (2000 gpm @ 28 ft H2O)	1	1
9	Hydrobins	--	227 lpm (60 gpm)	1	1
10	Baghouse Hopper (part of baghouse scope of supply)	--	--	24	0
11	Air Heater Hopper (part of boiler scope of supply)	--	--	10	0
12	Air Blower	--	20 m3/min @ 0.2 MPa (710 scfm @ 24 psi)	1	1
13	Fly Ash Silo	Reinforced concrete	635 tonne (1,400 ton)	2	0
14	Slide Gate Valves	--	--	2	0
15	Unloader	--	--	1	0
16	Telescoping Unloading Chute	--	127 tonne/h (140 tph)	1	0

Account 11 – Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	STG Transformer	Oil-filled	24 kV/345 kV, 650 MVA, 3-ph, 60 Hz	1	0
2	Auxiliary Transformer	Oil-filled	24 kV/4.16 kV, 250 MVA, 3-ph, 60 Hz	1	1
3	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 37 MVA, 3-ph, 60 Hz	1	1
4	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
5	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
6	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
7	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Account 12 – Instrumentation and Control

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	DCS - Main Control	Monitor/keyboard; Operator printer (laser color); Engineering printer (laser B&W)	Operator stations/printers and engineering stations/printers	1	0
2	DCS - Processor	Microprocessor with redundant input/output	N/A	1	0
3	DCS - Data Highway	Fiber optic	Fully redundant, 25% spare	1	0

7. COST ESTIMATING RESULTS

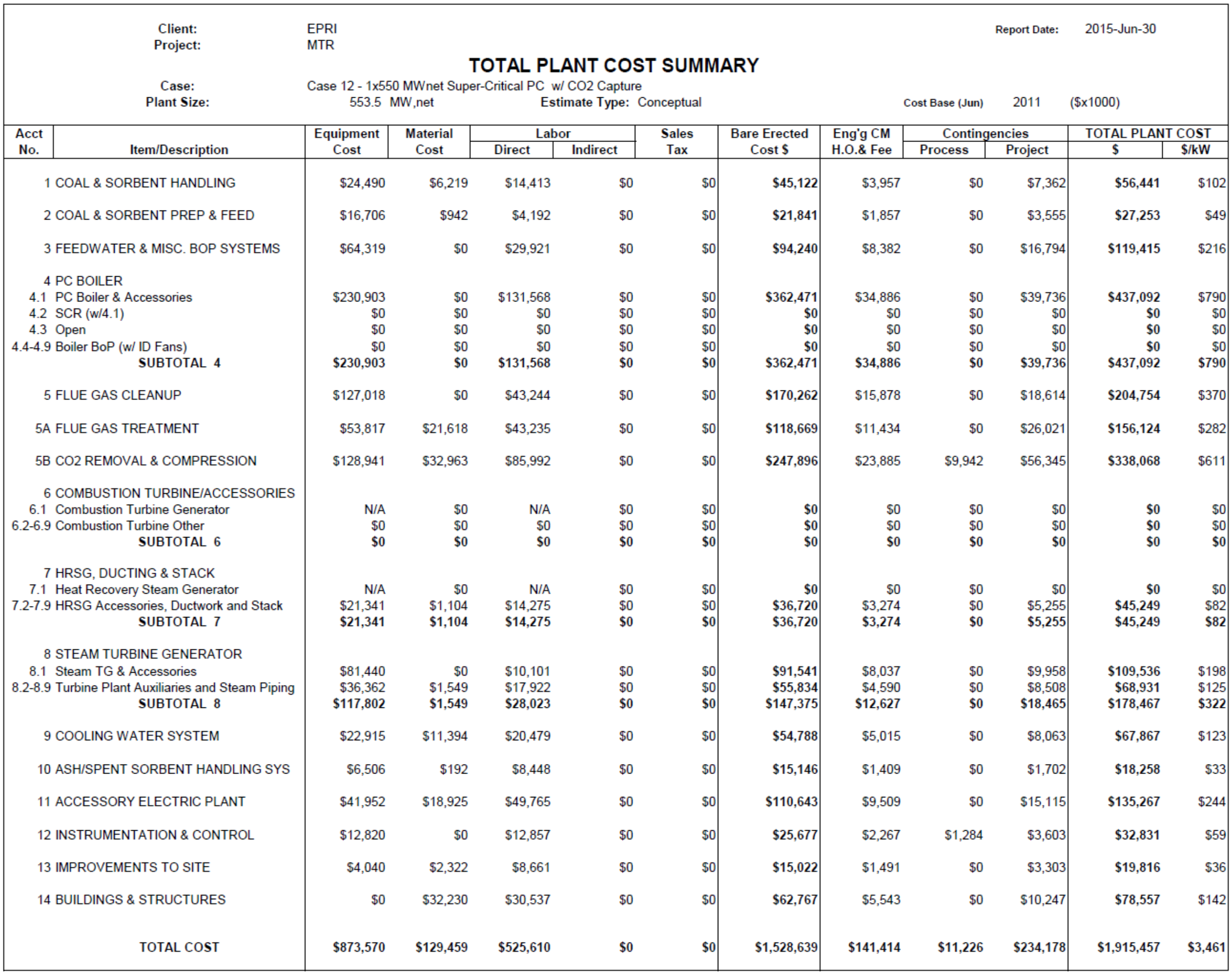
The cost estimating methodology is described in Section 2.7 of the BBS report [1], with updates presented in reference [2]. One notable cost adjustment was made to the boiler. Costs for the boiler were increased by 10% to account for changes required to accommodate the MTR system. Changes include: reduction in oxygen in the secondary combustion air, increased secondary air flow, reduced furnace temperature, increased furnace heat transfer area, increased convection pass cross sectional flow area to maintain the velocity design criteria to avoid excessive erosion. Given the current level of maturity of the MTR technology (small pilot scale), a process contingency of 20% was applied to the MTR CO_2 membrane modules and 15% was applied to the CO_2 purification system capital costs. A maturity level of the post-combustion CO_2 removal system in the BBS report was judged to be "process unproven at commercial scale for power plant applications, full-sized modules have been operated", and a 20% process contingency was applied to the post combustion CO_2 capture system capital costs in BBS report.

Exhibit 7-1 shows a cost summary of the design cases equipped with MTR CO_2 capture system in comparison to BBS Cases 11 and 12. Total plant capital costs for the MTR design cases organized by cost account are presented in Exhibit 7-2 and Exhibit 7-3. Initial and annual O&M costs are provided in Exhibit 7-4.

Exhibit 7-1:
Comparative Summary of Capital and Operating Costs (June, 2011USD, $x1000)

Item	BBS Case 11	BBS Case 12	MTR Case
Capital Costs			
Total Plant Costs (TPC)	$1,089,771	$1,959,399	$1,915,457
Annual Operating Costs			
Fixed Costs	$38,829	$64,138	$62,858
Variable Costs	$31,688	$54,089	$46,634
Fuel Costs	$104,591	$144,504	$144,542

Exhibit 7-2: MTR Case Total Plant Cost Summary



Client: EPRI
Project: MTR
Report Date: 2015-Jun-30

TOTAL PLANT COST SUMMARY

Case: Case 12 - 1x550 MWnet Super-Critical PC w/ CO2 Capture
Plant Size: 553.5 MW,net
Estimate Type: Conceptual
Cost Base (Jun) 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor		Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies		TOTAL PLANT COST	
				Direct	Indirect				Process	Project	$	$/kW
1	COAL & SORBENT HANDLING	$24,490	$6,219	$14,413	$0	$0	**$45,122**	$3,957	$0	$7,362	**$56,441**	$102
2	COAL & SORBENT PREP & FEED	$16,706	$942	$4,192	$0	$0	**$21,841**	$1,857	$0	$3,555	**$27,253**	$49
3	FEEDWATER & MISC. BOP SYSTEMS	$64,319	$0	$29,921	$0	$0	**$94,240**	$8,382	$0	$16,794	**$119,415**	$216
4	PC BOILER											
4.1	PC Boiler & Accessories	$230,903	$0	$131,568	$0	$0	**$362,471**	$34,886	$0	$39,736	**$437,092**	$790
4.2	SCR (w/4.1)	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
4.3	Open	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
4.4-4.9	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
	SUBTOTAL 4	**$230,903**	**$0**	**$131,568**	**$0**	**$0**	**$362,471**	**$34,886**	**$0**	**$39,736**	**$437,092**	**$790**
5	FLUE GAS CLEANUP	$127,018	$0	$43,244	$0	$0	**$170,262**	$15,878	$0	$18,614	**$204,754**	$370
5A	FLUE GAS TREATMENT	$53,817	$21,618	$43,235	$0	$0	**$118,669**	$11,434	$0	$26,021	**$156,124**	$282
5B	CO2 REMOVAL & COMPRESSION	$128,941	$32,963	$85,992	$0	$0	**$247,896**	$23,885	$9,942	$56,345	**$338,068**	$611
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	**$0**	$0	$0	$0	**$0**	$0
6.2-6.9	Combustion Turbine Other	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
	SUBTOTAL 6	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	**$0**	$0	$0	$0	**$0**	$0
7.2-7.9	HRSG Accessories, Ductwork and Stack	$21,341	$1,104	$14,275	$0	$0	**$36,720**	$3,274	$0	$5,255	**$45,249**	$82
	SUBTOTAL 7	**$21,341**	**$1,104**	**$14,275**	**$0**	**$0**	**$36,720**	**$3,274**	**$0**	**$5,255**	**$45,249**	**$82**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$81,440	$0	$10,101	$0	$0	**$91,541**	$8,037	$0	$9,958	**$109,536**	$198
8.2-8.9	Turbine Plant Auxiliaries and Steam Piping	$36,362	$1,549	$17,922	$0	$0	**$55,834**	$4,590	$0	$8,508	**$68,931**	$125
	SUBTOTAL 8	**$117,802**	**$1,549**	**$28,023**	**$0**	**$0**	**$147,375**	**$12,627**	**$0**	**$18,465**	**$178,467**	**$322**
9	COOLING WATER SYSTEM	$22,915	$11,394	$20,479	$0	$0	**$54,788**	$5,015	$0	$8,063	**$67,867**	$123
10	ASH/SPENT SORBENT HANDLING SYS	$6,506	$192	$8,448	$0	$0	**$15,146**	$1,409	$0	$1,702	**$18,258**	$33
11	ACCESSORY ELECTRIC PLANT	$41,952	$18,925	$49,765	$0	$0	**$110,643**	$9,509	$0	$15,115	**$135,267**	$244
12	INSTRUMENTATION & CONTROL	$12,820	$0	$12,857	$0	$0	**$25,677**	$2,267	$1,284	$3,603	**$32,831**	$59
13	IMPROVEMENTS TO SITE	$4,040	$2,322	$8,661	$0	$0	**$15,022**	$1,491	$0	$3,303	**$19,816**	$36
14	BUILDINGS & STRUCTURES	$0	$32,230	$30,537	$0	$0	**$62,767**	$5,543	$0	$10,247	**$78,557**	$142
	TOTAL COST	**$873,570**	**$129,459**	**$525,610**	**$0**	**$0**	**$1,528,639**	**$141,414**	**$11,226**	**$234,178**	**$1,915,457**	**$3,461**

Exhibit 7-3: MTR Case Total Plant Cost Details

Client: EPRI
Project: MTR
Report Date: 2015-Jun-30

TOTAL PLANT COST SUMMARY

Case: Case 12 - 1x550 MWnet Super-Critical PC w/ CO2 Capture
Plant Size: 553.5 MW,net **Estimate Type:** Conceptual **Cost Base (Jun)** 2011 ($x1000)

Acct		Equipment	Material	Labor		Sales	Bare Erected	Eng'g CM	Contingencies		TOTAL PLANT COST	
No.	Item/Description	Cost	Cost	Direct	Indirect	Tax	Cost $	H.O.& Fee	Process	Project	$	$/kW
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$4,996	$0	$2,251	$0	$0	$7,247	$628	$0	$1,181	$9,056	$16
1.2	Coal Stackout & Reclaim	$6,457	$0	$1,443	$0	$0	$7,900	$669	$0	$1,285	$9,854	$18
1.3	Coal Conveyors	$6,003	$0	$1,428	$0	$0	$7,431	$631	$0	$1,209	$9,270	$17
1.4	Other Coal Handling	$1,571	$0	$330	$0	$0	$1,901	$161	$0	$309	$2,371	$4
1.5	Sorbent Receive & Unload	$207	$0	$62	$0	$0	$269	$23	$0	$44	$335	$1
1.6	Sorbent Stackout & Reclaim	$3,343	$0	$604	$0	$0	$3,947	$333	$0	$642	$4,922	$9
1.7	Sorbent Conveyors	$1,193	$259	$289	$0	$0	$1,740	$146	$0	$283	$2,169	$4
1.8	Other Sorbent Handling	$721	$170	$373	$0	$0	$1,263	$108	$0	$206	$1,577	$3
1.9	Coal & Sorbent Hnd.Foundations	$0	$5,790	$7,635	$0	$0	$13,425	$1,259	$0	$2,203	$16,886	$31
	SUBTOTAL 1.	**$24,490**	**$6,219**	**$14,413**	**$0**	**$0**	**$45,122**	**$3,957**	**$0**	**$7,362**	**$56,441**	**$102**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$2,899	$0	$557	$0	$0	$3,456	$292	$0	$562	$4,310	$8
2.2	Coal Conveyor to Storage	$7,422	$0	$1,598	$0	$0	$9,020	$763	$0	$1,467	$11,251	$20
2.3	Coal Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.4	Misc.Coal Prep & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.5	Sorbent Prep Equipment	$5,699	$247	$1,167	$0	$0	$7,113	$599	$0	$1,157	$8,869	$16
2.6	Sorbent Storage & Feed	$687	$0	$259	$0	$0	$946	$81	$0	$154	$1,182	$2
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$696	$610	$0	$0	$1,306	$121	$0	$214	$1,641	$3
	SUBTOTAL 2.	**$16,706**	**$942**	**$4,192**	**$0**	**$0**	**$21,841**	**$1,857**	**$0**	**$3,555**	**$27,253**	**$49**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	Feedwater System	$26,722	$0	$8,616	$0	$0	$35,338	$3,020	$0	$5,754	$44,111	$80
3.2	Water Makeup & Pretreating	$7,759	$0	$2,454	$0	$0	$10,213	$933	$0	$2,229	$13,375	$24
3.3	Other Feedwater Subsystems	$8,406	$0	$3,451	$0	$0	$11,857	$1,019	$0	$1,931	$14,808	$27
3.4	Service Water Systems	$1,554	$0	$813	$0	$0	$2,367	$213	$0	$516	$3,095	$6
3.5	Other Boiler Plant Systems	$10,582	$0	$10,005	$0	$0	$20,587	$1,873	$0	$3,369	$25,829	$47
3.6	FO Supply Sys & Nat Gas	$390	$0	$455	$0	$0	$844	$76	$0	$138	$1,058	$2
3.7	Waste Treatment Equipment	$5,090	$0	$2,947	$0	$0	$8,037	$774	$0	$1,762	$10,572	$19
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$3,816	$0	$1,180	$0	$0	$4,997	$475	$0	$1,094	$6,566	$12
	SUBTOTAL 3.	**$64,319**	**$0**	**$29,921**	**$0**	**$0**	**$94,240**	**$8,382**	**$0**	**$16,794**	**$119,415**	**$216**
4	PC BOILER											
4.1	PC Boiler & Accessories	$230,903	$0	$131,568	$0	$0	$362,471	$34,886	$0	$39,736	$437,092	$790
4.2	SCR (w/4.1)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Secondary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.8	Major Component Rigging	$0	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations	$0	w/14.1	w/14.1	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$230,903**	**$0**	**$131,568**	**$0**	**$0**	**$362,471**	**$34,886**	**$0**	**$39,736**	**$437,092**	**$790**
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$87,501	$0	$18,708	$0	$0	$106,209	$9,860	$0	$11,607	$127,676	$231
5.2	Other FGD	$4,566	$0	$5,139	$0	$0	$9,705	$920	$0	$1,062	$11,687	$21
5.3	Bag House & Accessories	$25,858	$0	$16,298	$0	$0	$42,155	$3,961	$0	$4,612	$50,729	$92
5.4	Other Particulate Removal Materials	$1,750	$0	$1,860	$0	$0	$3,610	$342	$0	$395	$4,346	$8
5.5	Gypsum Dewatering System	$7,343	$0	$1,239	$0	$0	$8,582	$795	$0	$938	$10,315	$19
5.6	Mercury Removal System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
5.9	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 5.	**$127,018**	**$0**	**$43,244**	**$0**	**$0**	**$170,262**	**$15,878**	**$0**	**$18,614**	**$204,754**	**$370**
5A	FLUE GAS TREATMENT											
5A.1	In-situ Flue Gas / Glycol Heat HX	$3,135	$1,359	$2,717	$0	$0	$7,212	$695	$0	$1,581	$9,488	$17
5A.2	Flue Gas Compressor	$25,170	$10,068	$20,136	$0	$0	$55,374	$5,335	$0	$12,142	$72,851	$132
5A.3	Direct Contact Cooler	$2,833	$1,180	$2,361	$0	$0	$6,375	$614	$0	$1,398	$8,386	$15
5A.4	In-situ Glycol / Flue Gas Heat HX (HX1-C)	$4,374	$1,895	$3,791	$0	$0	$10,060	$969	$0	$2,206	$13,235	$24
5A.5	Glycol Circulation Pumps	$245	$245	$490	$0	$0	$981	$94	$0	$215	$1,290	$2
5A.6	Flue Gas Turbo Expander	$16,291	$6,516	$13,033	$0	$0	$35,840	$3,453	$0	$7,859	$47,151	$85
5A.9	Booster Air Fan	$1,768	$354	$707	$0	$0	$2,829	$273	$0	$620	$3,722	$7
	SUBTOTAL 5A.	**$53,817**	**$21,618**	**$43,235**	**$0**	**$0**	**$118,669**	**$11,434**	**$0**	**$26,021**	**$156,124**	**$282**
5B	CO2 REMOVAL & COMPRESSION											
5B.1	MTR CO2 Membrane Modules	$15,400	$0	$15,400	$0	$0	$30,800	$2,968	$6,160	$7,986	$47,913	$87
5B.2	Compression Systems	$69,681	$20,741	$41,482	$0	$0	$131,904	$12,709	$0	$28,923	$173,536	$314
5B.3	CO2 Purification Systems	$17,904	$2,436	$4,873	$0	$0	$25,213	$2,429	$3,782	$6,285	$37,710	$68
5B.4	CO2 Chilling System	$19,577	$7,827	$15,653	$0	$0	$43,057	$4,149	$0	$9,441	$56,647	$102
5B.5	Ductwork to FD Fan	$6,378	$1,959	$8,585	$0	$0	$16,922	$1,630	$0	$3,710	$22,262	$40
	SUBTOTAL 5.	**$128,941**	**$32,963**	**$85,992**	**$0**	**$0**	**$247,896**	**$23,885**	**$9,942**	**$56,345**	**$338,068**	**$611**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$11,208	$0	$7,075	$0	$0	$18,283	$1,544	$0	$2,974	$22,802	$41
7.4	Stack	$10,133	$0	$5,888	$0	$0	$16,021	$1,504	$0	$1,752	$19,277	$35
7.9	Duct & Stack Foundations	$0	$1,104	$1,312	$0	$0	$2,416	$226	$0	$528	$3,170	$6
	SUBTOTAL 7.	**$21,341**	**$1,104**	**$14,275**	**$0**	**$0**	**$36,720**	**$3,274**	**$0**	**$5,255**	**$45,249**	**$82**

Exhibit 7-3: MTR Case Total Plant Cost Details (Cont'd)

Client: EPRI
Project: MTR
Report Date: 2015-Jun-30

TOTAL PLANT COST SUMMARY

Case: Case 12 - 1x550 MWnet Super-Critical PC w/ CO2 Capture
Plant Size: 553.5 MW,net
Estimate Type: Conceptual
Cost Base (Jun) 2011 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $	TOTAL PLANT COST $/kW
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$81,440	$0	$10,101	$0	$0	$91,541	$8,037	$0	$9,958	$109,536	$198
8.2	Turbine Plant Auxiliaries	$520	$0	$1,106	$0	$0	$1,625	$155	$0	$178	$1,959	$4
8.3	Condenser & Auxiliaries	$9,918	$0	$2,740	$0	$0	$12,658	$1,178	$0	$1,384	$15,219	$27
8.4	Steam Piping	$25,925	$0	$11,518	$0	$0	$37,443	$2,870	$0	$6,047	$46,360	$84
8.9	TG Foundations	$0	$1,549	$2,558	$0	$0	$4,108	$387	$0	$899	$5,394	$10
	SUBTOTAL 8.	**$117,802**	**$1,549**	**$28,023**	**$0**	**$0**	**$147,375**	**$12,627**	**$0**	**$18,465**	**$178,467**	**$322**
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$17,140	$0	$5,301	$0	$0	$22,442	$2,091	$0	$2,453	$26,985	$49
9.2	Circulating Water Pumps	$3,432	$0	$263	$0	$0	$3,696	$315	$0	$401	$4,412	$8
9.3	Circ.Water System Auxiliaries	$884	$0	$117	$0	$0	$1,001	$93	$0	$109	$1,203	$2
9.4	Circ.Water Piping	$0	$7,449	$6,746	$0	$0	$14,194	$1,256	$0	$2,318	$17,768	$32
9.5	Make-up Water System	$737	$0	$947	$0	$0	$1,684	$155	$0	$276	$2,116	$4
9.6	Component Cooling Water Sys	$721	$0	$553	$0	$0	$1,274	$116	$0	$209	$1,599	$3
9.9	Circ.Water System Foundations & Structure	$0	$3,945	$6,552	$0	$0	$10,497	$989	$0	$2,297	$13,783	$25
	SUBTOTAL 9.	**$22,915**	**$11,394**	**$20,479**	**$0**	**$0**	**$54,788**	**$5,015**	**$0**	**$8,063**	**$67,867**	**$123**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Ash Coolers	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.2	Cyclone Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.3	HGCU Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Recovery Equipment	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$851	$0	$2,605	$0	$0	$3,456	$332	$0	$379	$4,167	$8
10.7	Ash Transport & Feed Equipment	$5,655	$0	$5,606	$0	$0	$11,261	$1,037	$0	$1,230	$13,528	$24
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$192	$237	$0	$0	$429	$40	$0	$94	$563	$1
	SUBTOTAL 10.	**$6,506**	**$192**	**$8,448**	**$0**	**$0**	**$15,146**	**$1,409**	**$0**	**$1,702**	**$18,258**	**$33**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$2,312	$0	$370	$0	$0	$2,681	$240	$0	$219	$3,141	$6
11.2	Station Service Equipment	$8,086	$0	$2,711	$0	$0	$10,797	$1,003	$0	$885	$12,686	$23
11.3	Switchgear & Motor Control	$9,281	$0	$1,612	$0	$0	$10,894	$1,007	$0	$1,190	$13,091	$24
11.4	Conduit & Cable Tray	$0	$6,365	$20,563	$0	$0	$26,928	$2,513	$0	$4,416	$33,857	$61
11.5	Wire & Cable	$0	$12,119	$21,663	$0	$0	$33,782	$2,713	$0	$5,474	$41,969	$76
11.6	Protective Equipment	$412	$0	$1,429	$0	$0	$1,841	$177	$0	$202	$2,220	$4
11.7	Standby Equipment	$1,725	$0	$40	$0	$0	$1,765	$162	$0	$193	$2,119	$4
11.8	Main Power Transformers	$20,136	$0	$253	$0	$0	$20,390	$1,545	$0	$2,193	$24,128	$44
11.9	Electrical Foundations	$0	$442	$1,124	$0	$0	$1,566	$148	$0	$343	$2,057	$4
	SUBTOTAL 11.	**$41,952**	**$18,925**	**$49,765**	**$0**	**$0**	**$110,643**	**$9,509**	**$0**	**$15,115**	**$135,267**	**$244**
12	INSTRUMENTATION & CONTROL											
12.1	PC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$640	$0	$391	$0	$0	$1,031	$97	$52	$177	$1,356	$2
12.7	Distributed Control System Equipment	$6,459	$0	$1,152	$0	$0	$7,611	$703	$381	$869	$9,564	$17
12.8	Instrument Wiring & Tubing	$3,895	$0	$7,088	$0	$0	$10,983	$890	$549	$1,863	$14,285	$26
12.9	Other I & C Equipment	$1,825	$0	$4,226	$0	$0	$6,052	$578	$303	$693	$7,626	$14
	SUBTOTAL 12.	**$12,820**	**$0**	**$12,857**	**$0**	**$0**	**$25,677**	**$2,267**	**$1,284**	**$3,603**	**$32,831**	**$59**
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$68	$1,445	$0	$0	$1,513	$147	$0	$332	$1,992	$4
13.2	Site Improvements	$0	$2,254	$2,978	$0	$0	$5,232	$519	$0	$1,150	$6,902	$12
13.3	Site Facilities	$4,040	$0	$4,238	$0	$0	$8,277	$824	$0	$1,820	$10,921	$20
	SUBTOTAL 13.	**$4,040**	**$2,322**	**$8,661**	**$0**	**$0**	**$15,022**	**$1,491**	**$0**	**$3,303**	**$19,816**	**$36**
14	BUILDINGS & STRUCTURES											
14.1	Boiler Building	$0	$11,512	$10,117	$0	$0	$21,629	$1,903	$0	$3,530	$27,062	$49
14.2	Turbine Building	$0	$16,924	$15,762	$0	$0	$32,686	$2,884	$0	$5,335	$40,905	$74
14.3	Administration Building	$0	$814	$860	$0	$0	$1,673	$149	$0	$273	$2,095	$4
14.4	Circulation Water Pumphouse	$0	$222	$176	$0	$0	$398	$35	$0	$65	$498	$1
14.5	Water Treatment Buildings	$0	$968	$882	$0	$0	$1,849	$163	$0	$302	$2,314	$4
14.6	Machine Shop	$0	$544	$365	$0	$0	$909	$79	$0	$148	$1,137	$2
14.7	Warehouse	$0	$369	$370	$0	$0	$738	$65	$0	$121	$924	$2
14.8	Other Buildings & Structures	$0	$301	$256	$0	$0	$558	$49	$0	$91	$698	$1
14.9	Waste Treating Building & Str.	$0	$577	$1,749	$0	$0	$2,326	$216	$0	$381	$2,924	$5
	SUBTOTAL 14.	**$0**	**$32,230**	**$30,537**	**$0**	**$0**	**$62,767**	**$5,543**	**$0**	**$10,247**	**$78,557**	**$142**
	TOTAL COST	**$873,570**	**$129,459**	**$525,610**	**$0**	**$0**	**$1,528,639**	**$141,414**	**$11,226**	**$234,178**	**$1,915,457**	**$3,461**

Exhibit 7-4: MTR Case Initial and Annual Operating and Maintenance Costs

INITIAL & ANNUAL O&M EXPENSES

Case 12 - 1x550 MWnet Super-Critical PC w/ CO2 Capture

Cost Base (Jun):	2011
Heat Rate-net (Btu/kWh):	12,004
MWe-net:	553.5
Capacity Factor (%):	85

OPERATING & MAINTENANCE LABOR

Operating Labor

Operating Labor Rate (base):	39.70	$/hour
Operating Labor Burden:	30.00	% of base
Labor O-H Charge Rate:	25.00	% of labor

		Total
Skilled Operator	2.0	2.0
Operator	11.3	11.3
Foreman	1.0	1.0
Lab Tech's, etc.	2.0	2.0
TOTAL-O.J.'s	16.3	16.3

	Annual Cost $	Annual Unit Cost $/kW-net
Annual Operating Labor Cost	$7,384,208	$13.426
Maintenance Labor Cost	$12,254,876	$22.282
Administrative & Support Labor	$4,909,771	$8.927
Property Taxes and Insurance	$38,309,136	$69.653
TOTAL FIXED OPERATING COSTS	**$62,857,991**	**$114.287**

VARIABLE OPERATING COSTS

		$/kWh-net
Maintenance Material Cost	**$18,382,314**	**$0.00449**

Consumables	Consumption Initial Fill	Consumption /Day	Unit Cost	Initial Fill Cost		
Water(/1000 gallons)	0	6,414	1.67	**$0**	**$3,331,029**	**$0.00081**
Chemicals						
MU & WT Chem.(lbs)	0	31,047	0.27	$0	$2,579,890	$0.00063
Limestone (ton)	0	705	33.48	$0	$7,322,307	$0.00179
Carbon (Mercury Removal) lb	0	0	1.63	$0	$0	$0.00000
Replacement Membrane Modules (m2)	0	562.56	10.00	$0	$1,745,333	$0.00043
NaOH (tons)	0	0.00	671.16	$0	$0	$0.00000
H2SO4 (tons)	0	0.00	214.78	$0	$0	$0.00000
Corrosion Inhibitor	0	0	0.00	$0	$0	$0.00000
Activated Carbon (lb)	0	0	1.63	$0	$0	$0.00000
Ammonia (19% NH3) ton	0	68	330.00	$0	$6,962,010	$0.00170
Subtotal Chemicals				**$0**	**$18,609,540**	**$0.00454**
Other						
Supplemental Fuel (MBtu)	0	0	0.00	$0	$0	$0.00000
SCR Catalyst (m3)	w/equip.	0.43	8,938.80	$0	$1,178,637	$0.00029
Emission Penalties	0	0	0.00	$0	$0	$0.00000
Subtotal Other				**$0**	**$1,178,637**	**$0.00029**
Waste Disposal						
Fly Ash (ton)	0	527	25.11	$0	$4,105,733	$0.00100
Bottom Ash (ton)	0	132	25.11	$0	$1,026,457	$0.00025
Subtotal-Waste Disposal				**$0**	**$5,132,190**	**$0.00125**
By-products & Emissions						
Gypsum (tons)	0	1,213	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$0**	**$46,633,711**	**$0.01139**
Fuel (ton)	0	6,791	68.60	$0	**$144,542,385**	**$0.03529**

8. SUMMARY OF RESULTS AND FUTURE RECOMMENDATIONS

8.1 Results Summary

The performance and cost results of the MTR design case in comparison to BBS Cases 11 and 12 are summarized in Exhibit 8-1.

Exhibit 8-1: Performance and Cost Results

	BBS Case 11	BBS Case 12	BBS Case 11 with MTR Capture
Gross Power Output, kWe	580,400	662,800	780,795
Auxiliary Power Requirements, kWe	30,410	112,830	227,303
Net Power Output, kWe	549,990	549,970	553,492
HHV Thermal Input, kWth (MMBtu/h)	1,400,162 (4,778)	1,934,519 (6,601)	1,934,964 (6,602)
Net Plant HHV Efficiency, %	39.3	28.4	28.6
Net Plant HHV Efficiency Penalty, % Point	0.0	10.9	10.7
Raw Water Withdrawal, m^3/MWh Net (gpm/MWh net)	2.2 (9.7)	4.2 (18.3)	3.6 (15.8)
Raw Water Consumption, m^3/MWh Net (gpm/MWh net)	1.7 (7.7)	3.2 (14.1)	2.6 (11.4)
CO_2 Generated, kg/h (lb/h)[A]	440,322 (970,728)	608,090 (1,340,587)	608,438 (1,341,353)
Capture Efficiency, %	0.0	90.2	89.9
CO_2 Emitted, kg/h (lb/h)	440,322 (970,728)	59,697 (131,608)	61,239 (135,007)
CO_2 Emissions, kg/MWh Gross (lb/MWh Gross)	759 (1,673)	90 (199)	78 (173)
CO_2 Emissions, kg/MWh Net (lb/MWh Gross)	801 (1,765)	109 (239)	111 (244)
Total Plant Cost, Jun 2011$ x 1,000	1,089,771	1,959,399	1,915,457
Change in TPC, %	N/A	80	75

Notes:

B. Based on 100% carbon conversion.

The major findings of this preliminary TEA are as follows:

1. As compared to BBS Case 11, the addition of the MTR CO_2 membrane capture system necessitated an increase in plant gross power output by 200.4 MWe. While for BBS Case 12, power output increased by 82.4 MWe.
2. The penalty in net plant efficiency for the MTR case is 10.7 percentage points. In comparison to BBS Case 12 (10.9 percentage points penalty), the MTR configuration demonstrated a small relative improvement in net plant efficiency penalty.
3. The addition of MTR CO_2 membrane capture system resulted in an increase of TPC (compared to BBS Case 11) by 75%. TPC penalty for the MTR case is an improvement in comparison to the BBS Case 12 TPC penalty of 80%.
4. The overall performance and cost results for MTR provide an improvement over the results of the reference BBS Case 12.

8.2 Recommendations for Future Consideration

Below is a subsection that document recommendations for future technical considerations.

8.2.1 Alternate Configuration for Lower CO_2 Capture

The base configuration presented in Exhibit 3-1 was developed to meet the 90% CO_2 removal requirement with the MTR CO_2 membrane capture system. This 90% capture requirement mandated the use of counter-flow modules (Module B) in addition to the cross-flow modules (Module A). The counter-flow modules provide a second step of CO_2 capture in order to meet the capture target. The counter-flow modules utilize sweep air to provide for the CO_2 partial pressure differential driving force for separation. This sweep air leaving the module is recycled back to the PC boiler for use as secondary combustion air.

Where less than 90% CO_2 capture is required, a membrane process that eliminates the counter-current sweep module could be entertained, thus eliminating the performance impact on the steam generator and secondary air system. If such a capture system were desired, this has the potential to be a significant sweet spot for the membrane system.

The current proposed EPA new source performance standard regulations on CO_2 emission in 40CFR60, Subpart Da (see 79Federal Register1430, dated January 8, 2014) are based on a 1,100 pounds (lb) of CO_2 per megawatt-hour (MWh) of gross energy output on a 12-operating month rolling average basis emission rate for new fossil plants. The BBS Case 11 reference has a CO_2 emission of 1673 lb/MWh, gross, which would require a CO_2 capture rate of approximately 40%.

The development of a configuration to support a 40 to 50% CO_2 capture rate could look like that presented in Exhibit 8-2**.** Benefits of this simplified configuration are presented in Exhibit 8-3.

Exhibit 8-2: Block Flow Diagram of MTR CO_2 Capture Process for 40-50% Capture

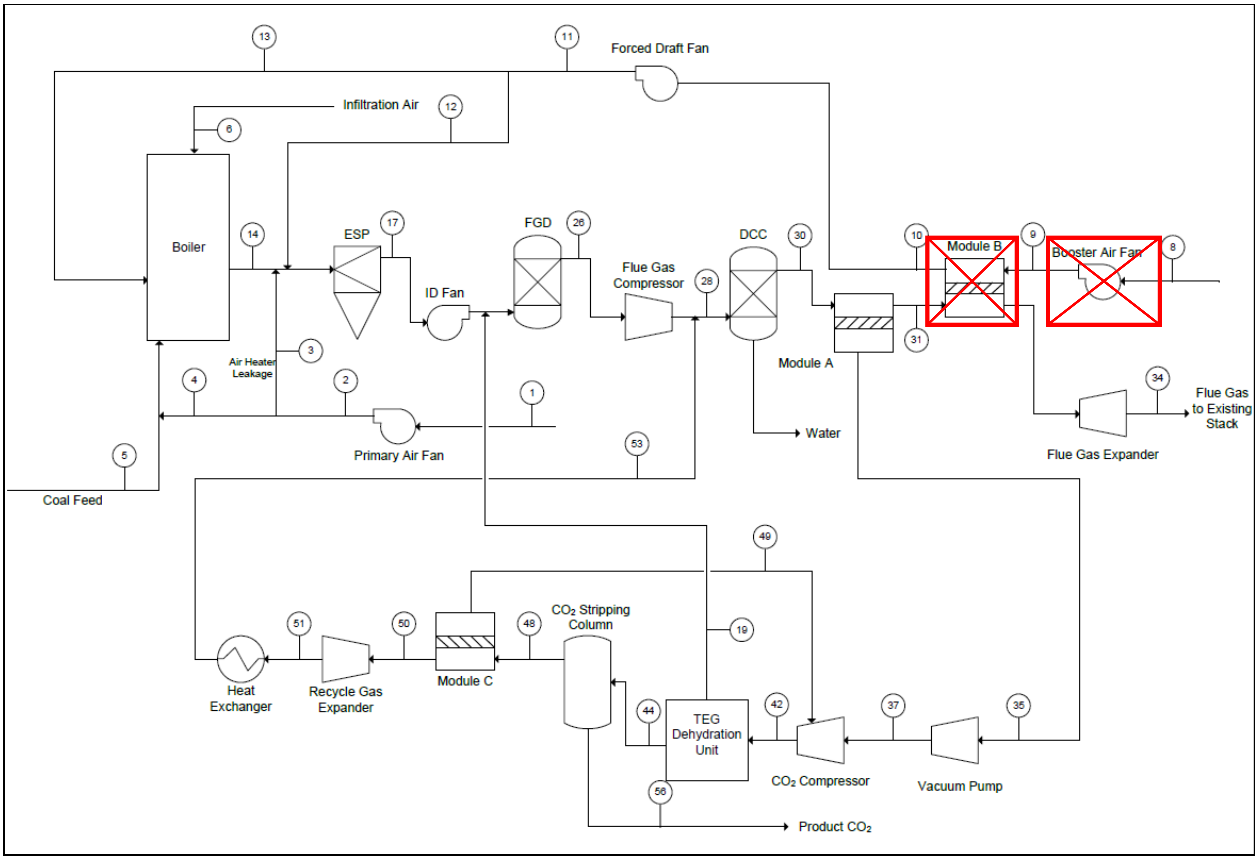


Exhibit 8-3: Advantages of a Simplified Configuration for a 40-50% CO_2 Capture

Change	Capital Benefit	Performance Benefit	Notes
Eliminate Module B	✓	✓	Eliminates pressure loss for booster air fan.
Eliminate Booster Air Fan	✓	✓	Eliminates booster air fan aux. load.
Minimize FD Fan	✓	✓	FD fan flow is reduced / head is unchanged.
Eliminate boiler design impact	✓	✓	Eliminate boiler cost penalty. Eliminate 2% boiler efficiency penalty
Eliminate baghouse design impact	✓	✓	Baghouse flow reduced as there will be no additional N_2 nor CO_2 from the air sweep. No additional PM loading from boiler efficiency penalty.
Minimize ID Fan	✓	✓	ID fan head unchanged. Flow reduced as no additional N_2/ CO_2 from air sweep.

Change	Capital Benefit	Performance Benefit	Notes
Minimize FGD	✓		Flow rate reduced from 90% capture MTR case, as now air sweep recycle.
Flue Gas Compressor	?	?	Flowrate is reduced with the elimination of the air sweep, however head may still need to be increased to provide better driving force across Module A in light of the reduced CO_2 concentration. It may be possible to eliminate flue gas compression for partial capture cases. This option is under investigation.
Minimized DCC	✓		Reduced flow rate and DCC size
Flue gas expander	✗	✓	Cost will be greater as pressure ratio and power output may be increased. Pressure ratio is greater from the elimination of Module B.
CO_2 compression, drying purification train	✓	✓	Less CO_2 to process makes whole system smaller, reducing cost and aux. load.
Minimize STG & cycle	✓	✓	The reduced aux. load will permit a smaller steam turbine, GSU, & condenser.
Minimize cooling tower and CO_2 capture cooling	✓	✓	The reduced STG and CO_2 flowrate will minimize cost and performance impact.
Minimize Refrigeration	✓	✓	The reduced CO_2 flowrate to be purified will reduce the refrigeration requirement.
OVERALL PROCESS	✓	✓	**Overall process improvements are expected to be significant.**

In addition to the above benefits and noted drivers, all equipment/systems will be smaller with reduced CO_2 capture, as there will be less auxiliary load. The gross electric generation requirement will be reduced, and all systems will be reduced in size accordingly. For simplicity and clarity, this is not listed above for the boiler flue gas processes in order to emphasize there are process benefits beyond the benefits simply related to the reduced quantity of CO_2 captured.

While these alternative configurations have not been evaluated in this technical assessment, it is recommended that MTR consider this configuration for future assessment.

As similar benefits would also accrue to the reference BBS Case 12 for a reduced carbon capture level, it is recommended that both cases be evaluated with nominal 40%-50% CO_2 capture. This will provide a “real world” reference for the comparison between conventional amine capture and the MTR system.

9. REFERENCES

1 Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2, November 2010, DOE/NETL-2010/1397, http://www.netl.doe.gov/research/energy-analysis/energy-baseline-studies

2 Updated Costs (June 2011 Basis) for Selected Bituminous Baseline Cases, August 2012, DOE/NETL-341/082312, http://www.netl.doe.gov/File%20Library/Research/Energy%20Analysis/Publications/BaselineCostUpdate.pdf

3 Email from Tim Merkel (MTR) to David Stauffer (WorleyParsons), Dated 2015-05-13, RE: MTR CO2 Capture Rate: 89% acceptable?, (includes MTR membrane H&MB).

4 Email from Tim Merkel (MTR) to David Stauffer (WorleyParsons), Dated 2010-03-08, Re: Mass Balances for the minimal and “Optimal” Feed Compression Cases.

5 Email from Tim Merkel (MTR) to David Stauffer (WorleyParsons), Dated 2010-05-06, Re: Membrane Feed Pressure Sensitivity Data.

6 Email from Xiaotong Wei (MTR) to David Stauffer (WorleyParsons), Dated 2010-04-26, Re: Particulate Filtration Requirements.

7 Email from Tim Merkel (MTR) to David Stauffer (WorleyParsons), Dated 2010-03-23, Re: Supplemental Questions – Regarding Mass Balance data.

8 Email from Tim Merkel (MTR) to David Stauffer (WorleyParsons), Dated 2010-04-21, Re: EPRI MTR CO2 Membrane System - Information Request.

9 “Membrane Process to Capture CO2 From Coal-Fired Power Plant Flue Gas,” Award Quarterly Progress Report October 1, 2009 to December 31, 2009, Dated January 29, 2010.

10 Email from Tim Merkel (MTR) to David Stauffer (WorleyParsons), Dated 2010-05-13, Re: Design Basis Module Dimensions.

11 Email from Brice Freeman (MTR) to David Stauffer (WorleyParsons), Dated 2015-06-05, RE: EPRI MTR TEA Support- Membrane Cost Input.

12 Email from Tim Merkel (MTR) to David Stauffer (WorleyParsons), Dated 2015-05-28, RE: EPRI MTR TEA Support- Membrane Cost Input.

13 Email from Brice Freeman (MTR) to David Stauffer (WorleyParsons), Dated 2015-05-13, RE: MTR CO2 Capture Rate: 89% acceptable?, (includes results from Trimeric study).

14 MTR Membrane Process for Post-Combustion CO_2 Capture: A Systems Design and Economic Analysis, 30, March, 2012, DOE Contract Number DE-FE0005795.

15 Quality Guidelines for Energy System Studies, CO2 Impurity Design Parameters, August 2013, DOE/NETL-341/011212, http://www.netl.doe.gov/research/energy-analysis/quality-guidelines-qgess

Appendix 1: Heat and Mass Balance Tables

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EPRI / MTR CO_2 MEMBRANE
DESIGN CASE

STREAM	1		2		3		4		5		6		7	
DESCRIPTION	Ambient Air to Primary Air Fan		Primary Air		Primary Air Leakage to Flue Gas		Primary Air to Coal Milling		Coal Feed		Infiltration Air		Bottom Ash	
Vapor & Liquid														
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	419	0.92	419	0.92	58	0.92	361	0.92	0	N/A	36	0.92	0	N/A
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
CO_2	15	0.03	15	0.03	2	0.03	13	0.03	0	N/A	1	0.03	0	N/A
H_2O	447	0.99	447	0.99	62	0.99	386	0.99	0	N/A	38	0.99	0	N/A
N_2	35,024	77.32	35,024	77.32	4,820	77.32	30,204	77.32	0	N/A	2,973	77.32	0	N/A
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
O_2	9,395	20.74	9,395	20.74	1,293	20.74	8,102	20.74	0	N/A	797	20.74	0	N/A
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
TOTAL	45,300	100.00	45,300	100	6,234	100	39,065	100	0	N/A	3,845	100	0	N/A
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	16,739	1.28	16,739	1.28	2,304	1.28	14,435	1.28	0	N/A	1,421	1.28	0	N/A
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
CO_2	654	0.05	654	0.05	90	0.05	564	0.05	0	N/A	56	0.05	0	N/A
H_2O	8,055	0.62	8,055	0.62	1,109	0.62	6,946	0.62	0	N/A	684	0.62	0	N/A
N_2	981,138	75.06	981,138	75.06	135,031	75.06	846,107	75.06	0	N/A	83,276	75.06	0	N/A
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
O_2	300,633	23.00	300,633	23.00	41,375	23.00	259,258	23.00	0	N/A	25,517	23.00	0	N/A
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
TOTAL	1,307,219	100	1,307,219	100	179,908	100	1,127,310	100	0	N/A	110,953	100	0	N/A
Solid														
Solid Components	**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		565,950		0		0	
Ash	0		0		0		0		0		0		10,980	
$CaCO_3$	0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0	
$CaSO_4\cdot 2H_2O$	0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0	
TOTAL	0		0		0		0		565,950		0		10,980	
All Phases														
All Phases														
Temperature, °F	59		77		77		469		59		59		707	
Pressure, psia	14.7		16.1		16.1		16.1		14.7		14.7		14.7	
Total Flow, lb/h	1,307,219		1,307,219		179,908		1,127,310		565,950		110,953		10,980	
Enthalpy, Btu/lb (V&L Only)	-42.0		-37.5		-37.5		58.1		N/A		-42.0		N/A	
Flow Rate, MMSCFD (V Only)	413		413		57		356		N/A		35		N/A	
Vapor Frac	1.00		1.00		1.00		1.00		N/A		1.00		N/A	
Density, lb/ft^3 (V&L Only)	0.076		0.081		0.081		0.047		N/A		0.076		N/A	
Average MW (V&L Only)	28.86		28.86		28.86		28.86		N/A		28.86		N/A	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 14.696 psia and 60 °F.
3. Enthalpies are referenced to the constituent elements in their standard states at 77 °F and 14.696 psia.

Rev A
11-Jun-15

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EPRI / MTR CO_2 MEMBRANE
DESIGN CASE

STREAM	8		9		10		11		12		13	
DESCRIPTION	Ambient Air to Secondary Air Booster Fan		Secondary Air to Module B		Vitiated Secondary Air to Forced Draft Fan		Secondary Air to Air Heater		Secondary Air Leakage to Flue Gas		Secondary Air to Boiler	
Vapor & Liquid												
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	1,415	0.92	1,415	0.92	1,429	0.86	1,429	0.86	42	0.86	1,387	0.86
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2	50	0.03	50	0.03	12,553	7.56	12,553	7.56	372	7.56	12,182	7.56
H_2O	1,510	0.99	1,510	0.99	1,972	1.19	1,972	1.19	58	1.19	1,914	1.19
N_2	118,259	77.32	118,259	77.32	119,468	71.99	119,468	71.99	3,538	71.99	115,930	71.99
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	1	0.00	1	0.00	0	0.00	1	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	31,723	20.74	31,723	20.74	30,534	18.40	30,534	18.40	904	18.40	29,630	18.40
SO_2	0	0.00	0	0.00	2	0.00	2	0.00	0	0.00	2	0.00
TOTAL	152,957	100.00	152,957	100.00	165,960	100.00	165,960	100.00	4,915	100.00	161,045	100.00
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	56,520	1.28	56,520	1.28	57,091	1.15	57,091	1.15	1,691	1.15	55,400	1.15
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2	2,208	0.05	2,208	0.05	552,475	11.12	552,475	11.12	16,363	11.12	536,112	11.12
H_2O	27,197	0.62	27,197	0.62	35,528	0.71	35,528	0.71	1,052	0.71	34,476	0.71
N_2	3,312,840	75.06	3,312,840	75.06	3,346,730	67.35	3,346,730	67.35	99,124	67.35	3,247,600	67.35
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	27	0.00	27	0.00	1	0.00	26	0.00
NO_2	0	0.00	0	0.00	18	0.00	18	0.00	1	0.00	18	0.00
O_2	1,015,090	23.00	1,015,090	23.00	977,047	19.66	977,047	19.66	28,938	19.66	948,109	19.66
SO_2	0	0.00	0	0.00	136	0.00	136	0.00	4	0.00	132	0.00
TOTAL	4,413,855	100.00	4,413,855	100.00	4,969,051	100.00	4,969,051	100.00	147,174	100.00	4,821,872	100.00
Solid												
Solid Components	**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		0		0	
Ash	0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0	
$CaSO_4 \cdot 2H_2O$	0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0	
All Phases												
All Phases												
Temperature, °F	59		75		94		95		95		429	
Pressure, psia	14.7		16.0		15.2		15.3		15.3		15.0	
Total Flow, lb/h	4,413,855		4,413,855		4,969,051		4,969,051		147,174		4,821,872	
Enthalpy, Btu/lb (V&L Only)	-42.0		-38.0		-464.8		-464.6		-464.6		-383.5	
Flow Rate, MMSCFD (V Only)	1,393		1,393		1,511		1,511		45		1,467	
Vapor Frac	1.00		1.00		1.00		1.00		1.00		1.00	
Density, lb/ft^3 (V&L Only)	0.076		0.080		0.077		0.077		0.077		0.047	
Average MW (V&L Only)	28.86		28.86		29.94		29.94		29.94		29.94	

STREAM	14		15		16		17		18		19	
DESCRIPTION	Flue Gas to Air Heater		Flue Gas to Fabric Filter		Fly Ash		Flue Gas to ID Fans		ID Fan Discharge		TEG Dehydration Unit Vent Recycle	
Vapor & Liquid												
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	1,784	0.83	1,884	0.83	0	N/A	1,884	0.83	1,884	0.83	0	0.02
C_3H_8	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.05
Cl_2	23	0.01	23	0.01	0	N/A	23	0.01	23	0.01	2	0.40
CO_2	42,234	19.61	42,608	18.81	0	N/A	42,608	18.81	42,608	18.81	543	93.91
H_2O	18,558	8.62	18,678	8.25	0	N/A	18,678	8.25	18,678	8.25	24	4.10
N_2	149,386	69.37	157,744	69.65	0	N/A	157,744	69.65	157,744	69.65	3	0.58
NH_3	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.00
NO	10	0.00	10	0.00	0	N/A	10	0.00	10	0.00	0	0.00
NO_2	1	0.00	1	0.00	0	N/A	1	0.00	1	0.00	1	0.16
O_2	2,896	1.34	5,093	2.25	0	N/A	5,093	2.25	5,093	2.25	1	0.12
SO_2	445	0.21	445	0.20	0	N/A	445	0.20	445	0.20	4	0.65
TOTAL	215,337	100.00	226,486	100.00	0	N/A	226,486	100.00	226,486	100.00	578	100.00
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	71,256	1.08	75,250	1.09	0	N/A	75,250	1.09	75,250	1.09	5	0.02
C_3H_8	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	41	0.17
Cl_2	1,640	0.02	1,640	0.02	0	N/A	1,640	0.02	1,640	0.02	165	0.66
CO_2	1,858,720	28.28	1,875,180	27.18	0	N/A	1,875,180	27.18	1,875,180	27.18	23,891	95.84
H_2O	334,333	5.09	336,494	4.88	0	N/A	336,494	4.88	336,494	4.88	427	1.71
N_2	4,184,810	63.67	4,418,970	64.05	0	N/A	4,418,970	64.05	4,418,970	64.05	94	0.38
NH_3	7	0.00	7	0.00	0	N/A	7	0.00	7	0.00	0	0.00
NO	290	0.00	291	0.00	0	N/A	291	0.00	291	0.00	0	0.00
NO_2	26	0.00	26	0.00	0	N/A	26	0.00	26	0.00	43	0.17
O_2	92,665	1.41	162,979	2.36	0	N/A	162,979	2.36	162,979	2.36	23	0.09
SO_2	28,512	0.43	28,516	0.41	0	N/A	28,516	0.41	28,516	0.41	239	0.96
TOTAL	6,572,258	100.00	6,899,353	100.00	0	N/A	6,899,353	100.00	6,899,353	100.00	24,929	100.00
Solid												
Solid Components	**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		0		0	
Ash	43,919		43,919		43,919		0		0		0	
$CaCO_3$	0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0	
$CaSO_4 \cdot 2H_2O$	0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0	
TOTAL	43,919		43,919		43,919		0		0		0	
All Phases												
All Phases												
Temperature, °F	713		412		412		412		434		95	
Pressure, psia	14.5		14.4		14.2		14.2		15.3		16.8	
Total Flow, lb/h	6,616,178		6,943,273		43,919		6,899,353		6,899,353		24,929	
Enthalpy, Btu/lb (V&L Only)	-1,226.2		-1,250.9		N/A		-1,250.9		-1,245.1		-3,801.7	
Flow Rate, MMSCFD (V Only)	1,961		2,063		N/A		2,063		2,063		5	
Vapor Frac	1.00		1.00		N/A		1.00		1.00		1.00	
Density, lb/ft^3 (V&L Only)	0.035		0.047		N/A		0.046		0.049		0.122	
Average MW (V&L Only)	30.52		30.46		N/A		30.46		30.46		43.13	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 14.696 psia and 60 °F.
3. Enthalpies are referenced to the constituent elements in their standard states at 77 °F and 14.696 psia.

Rev A
11-Jun-15

EPRI / MTR CO_2 MEMBRANE RETROFIT
DESIGN CASE

STREAM	20		21		22		23		24		25		26		27		28		29		30		31	
DESCRIPTION	Flue Gas to Flue Gas Glycol Cooler		Flue Gas to FGD		FGD Limestone Slurry		FGD Make-Up Water		FGD Oxidation Air		Gypsum		Desulfurized Flue Gas		Flue Gas from Compressor		Flue Gas to Direct-Contact Cooler		Direct Contact Cooler Blowdown Water		Flue Gas to Module A		Flue Gas to Module B	
Vapor & Liquid																								
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	1,884	0.83	1,884	0.83	0	0.00	0	0.00	21	0.92	0	0.00	1,904	0.77	1,904	0.77	1,998	0.77	0	0.00	1,998	0.88	1,904	1.04
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	25	0.01	25	0.01	0	0.00	0	0.00	0	0.00	0	0.00	25	0.01	25	0.01	27	0.01	0	0.00	27	0.01	0	0.00
CO_2	43,151	19.00	43,151	19.00	0	0.00	0	0.00	1	0.03	0	0.01	43,592	17.61	43,592	17.61	46,254	17.88	0	0.00	46,254	20.38	15,638	8.51
H_2O	18,702	8.24	18,702	8.24	7,549	100.00	30,580	100.00	22	0.99	590	99.99	37,227	15.04	37,227	15.04	37,227	14.39	31,790	100.00	5,437	2.40	1,440	0.78
N_2	157,748	69.47	157,748	69.47	0	0.00	0	0.00	1,722	77.32	0	0.00	159,470	64.41	159,470	64.41	167,331	64.68	0	0.00	167,331	73.74	159,467	86.76
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	10	0.00	10	0.00	0	0.00	0	0.00	0	0.00	0	0.00	10	0.00	10	0.00	11	0.00	0	0.00	11	0.00	10	0.01
NO_2	2	0.00	2	0.00	0	0.00	0	0.00	0	0.00	0	0.00	2	0.00	2	0.00	2	0.00	0	0.00	1	0.00	0	0.00
O_2	5,094	2.24	5,094	2.24	0	0.00	0	0.00	462	20.74	0	0.00	5,336	2.16	5,336	2.16	5,859	2.26	0	0.00	5,859	2.58	5,335	2.90
SO_2	449	0.20	449	0.20	0	0.00	0	0.00	0	0.00	0	0.00	9	0.00	9	0.00	9	0.00	0	0.00	9	0.00	2	0.00
TOTAL	227,065	100.00	227,065	100.00	7,549	100.00	30,580	100.00	2,227	100.00	590.06	100.00	247,575	100.00	247,575	100.00	258,719	100.00	31,791	100.00	226,928	100.00	183,797	100.00
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	75,255	1.09	75,255	1.09	0	0.00	0	0.00	823	1.28	0	0.00	76,078	1.04	76,078	1.04	79,824	1.04	0	0.00	79,824	1.13	76,072	1.40
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	41	0.00	41	0.00	0	0.00	0	0.00	0	0.00	0	0.00	41	0.00	41	0.00	41	0.00	0	0.00	41	0.00	0	0.00
Cl_2	1,805	0.03	1,805	0.03	0	0.00	0	0.00	0	0.00	0	0.00	1,805	0.02	1,805	0.02	1,930	0.03	0	0.00	1,929	0.03	0	0.00
CO_2	1,899,070	27.43	1,899,070	27.43	0	0.00	0	0.00	32	0.05	3	0.03	1,918,460	26.26	1,918,460	26.26	2,035,640	26.56	8	0.00	2,035,630	28.71	688,223	12.68
H_2O	336,921	4.87	336,921	4.87	135,999	100.00	550,901	100.00	396	0.62	10,629	99.97	670,656	9.18	670,656	9.18	670,658	8.75	572,713	100.00	97,945	1.38	25,949	0.48
N_2	4,419,060	63.82	4,419,060	63.82	0	0.00	0	0.00	48,241	75.06	0	0.00	4,467,300	61.15	4,467,300	61.15	4,687,510	61.16	0	0.00	4,687,510	66.10	4,467,220	82.29
NH_3	7	0.00	7	0.00	0	0.00	0	0.00	0	0.00	0	0.00	7	0.00	7	0.00	8	0.00	1	0.00	6	0.00	0	0.00
NO	291	0.00	291	0.00	0	0.00	0	0.00	0	0.00	0	0.00	291	0.00	291	0.00	319	0.00	0	0.00	319	0.00	291	0.01
NO_2	69	0.00	69	0.00	0	0.00	0	0.00	0	0.00	0	0.00	69	0.00	69	0.00	69	0.00	1	0.00	68	0.00	23	0.00
O_2	163,001	2.35	163,001	2.35	0	0.00	0	0.00	14,782	23.00	0	0.00	170,743	2.34	170,743	2.34	187,494	2.45	0	0.00	187,494	2.64	170,707	3.14
SO_2	28,755	0.42	28,755	0.42	0	0.00	0	0.00	0	0.00	0	0.00	565	0.01	565	0.01	566	0.01	0	0.00	566	0.01	150	0.00
TOTAL	6,924,276	100.00	6,924,276	100.00	135,999	100.00	550,901	100.00	64,274	100.00	10,632	100.00	7,306,016	100.00	7,306,016	100.00	7,664,059	100.00	572,724	100.00	7,091,333	100.00	5,428,635	100.00
Solid																								
Solid Components	**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		55,807		0		0		11,767		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		75,760		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0		0		0		0		0	
SiO_2	0		0		1,762		0		0		1,762		0		0		0		0		0		0	
Fe_2O_3	0		0		1,175		0		0		1,175		0		0		0		0		0		0	
TOTAL	0		0		58,745		0		0		90,464		0		0		0		0		0		0	
All Phases																								
All Phases																								
Temperature, °F	433		332		59		59		59		135		135		273		266		95		95		95	
Pressure, psia	15.3		15.2		20.0		50.0		14.7		14.7		14.7		29.4		29.4		29.0		29.0		28.3	
Total Flow, lb/h	6,924,276		6,924,276		194,744		550,901		64,274		101,096		7,306,016		7,306,016		7,664,059		572,724		7,091,333		5,428,635	
Enthalpy, Btu/lb (V&L Only)	-1,254.3		-1,280.1		-6,886.2		-6,886.1		-42.0		-6,803.7		-1,524.9		-1,489.3		-1,478.2		-6,847.2		-1,179.4		-510.8	
Flow Rate, MMSCFD (V Only)	2,068		2,068		N/A		N/A		20		N/A		2,255		2,255		2,356		N/A		2,067		1,674	
Vapor Frac	1.00		1.00		0.00		0.00		1.00		0.00		1.00		1.00		1.00		0.00		1.00		1.00	
Density, lb/ft^3 (V&L Only)	0.049		0.055		62.650		62.650		0.076		51.772		0.068		0.110		0.112		61.445		0.153		0.141	
Average MW (V&L Only)	30.49		30.49		18.02		18.02		28.86		18.02		29.51		29.51		29.62		18.02		31.25		29.54	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 14.696 psia and 60 °F.
3. Enthalpies are referenced to the constituent elements in their standard states at 77 °F and 14.696 psia.

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EPRI / MTR CO_2 MEMBRANE
DESIGN CASE

STREAM	32		33		34		35		36		37		38		39	
DESCRIPTION	Flue Gas to Flue Gas Glycol Heater		Flue Gas to Expander		Flue Gas to Stack		Module A Permeate		Condensed Water from Vacuum Compressor		Raw CO_2 to Cooler		Raw CO_2 to Knock Out Drum		Condensed Water from Knock Out Drum	
Vapor & Liquid																
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	1,890	1.11	1,890	1.11	1,890	1.11	94	0.22	0	0.00	94	0.23	94	0.23	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	27	0.06	0	0.00	27	0.07	27	0.07	0	0.00
CO_2	3,135	1.84	3,135	1.84	3,135	1.84	30,616	70.98	0	0.00	30,616	73.79	30,616	73.79	0	0.00
H_2O	978	0.57	978	0.57	978	0.57	3,996	9.27	1,643	100.00	2,353	5.67	2,353	5.67	642	100.00
N_2	158,257	92.66	158,257	92.66	158,257	92.66	7,864	18.23	0	0.00	7,864	18.95	7,864	18.95	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	9	0.01	9	0.01	9	0.01	1	0.00	0	0.00	1	0.00	1	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	1	0.00	0	0.00	1	0.00	1	0.00	0	0.00
O_2	6,524	3.82	6,524	3.82	6,524	3.82	525	1.22	0	0.00	525	1.26	525	1.26	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	6	0.02	0	0.00	6	0.02	6	0.02	0	0.00
TOTAL	170,793	100.00	170,793	100.00	170,793	100.00	43,131	100.00	1,643	100.00	41,488	100.00	41,488	100.00	642	100.00
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	75,501	1.55	75,501	1.55	75,501	1.55	3,751	0.23	0	0.00	3,751	0.23	3,751	0.23	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	41	0.00	0	0.00	41	0.00	41	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	1,929	0.12	0	0.00	1,929	0.12	1,929	0.12	0	0.00
CO_2	137,956	2.83	137,956	2.83	137,956	2.83	1,347,400	81.04	1	0.00	1,347,400	82.51	1,347,400	82.51	0	0.00
H_2O	17,619	0.36	17,619	0.36	17,619	0.36	71,996	4.33	29,598	100.00	42,398	2.60	42,398	2.60	11,569	100.00
N_2	4,433,330	90.97	4,433,330	90.97	4,433,330	90.97	220,293	13.25	0	0.00	220,293	13.49	220,293	13.49	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	6	0.00	0	0.00	6	0.00	6	0.00	0	0.00
NO	264	0.01	264	0.01	264	0.01	29	0.00	0	0.00	29	0.00	29	0.00	0	0.00
NO_2	5	0.00	5	0.00	5	0.00	45	0.00	0	0.00	45	0.00	45	0.00	0	0.00
O_2	208,753	4.28	208,753	4.28	208,753	4.28	16,787	1.01	0	0.00	16,787	1.03	16,787	1.03	0	0.00
SO_2	14	0.00	14	0.00	14	0.00	416	0.03	0	0.00	416	0.03	416	0.03	0	0.00
TOTAL	4,873,442	100.00	4,873,442	100.00	4,873,442	100.00	1,662,694	100.00	29,599	100.00	1,633,096	100.00	1,633,096	100.00	11,569	100.00
Solid																
Solid Components	**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0	
All Phases																
All Phases																
Temperature, °F	76		225		128		95		95		150		95		95	
Pressure, psia	27.6		27.5		14.7		2.9		2.9		16.9		16.6		16.2	
Total Flow, lb/h	4,873,442		4,873,442		4,873,442		1,662,694		29,599		1,633,096		1,633,096		11,569	
Enthalpy, Btu/lb (V&L Only)	-130.1		-93.5		-117.2		-3,361.7		-6,847.2		-3,306.3		-3,326.3		-6,847.7	
Flow Rate, MMSCFD (V Only)	1,555		1,555		1,555		393		N/A		378		372		N/A	
Vapor Frac	1.00		1.00		1.00		1.00		0.00		1.00		0.98		0.00	
Density, lb/ft^3 (V&L Only)	0.137		0.107		0.067		0.019		57.853		0.102		0.112		61.462	
Average MW (V&L Only)	28.53		28.53		28.53		38.55		18.02		39.36		39.36		18.02	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 14.696 psia and 60 °F.
3. Enthalpies are referenced to the constituent elements in their standard states at 77 °F and 14.696 psia.

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EPRI / MTR CO_2 MEMBRANE DESIGN CASE

STREAM	40		41		42		43		44		45		46		47		48		49	
DESCRIPTION	Raw CO_2 to Multi-Stage Compressor		Condensed Water from Compressor		Raw CO_2 to TEG Unit		TEG Dehydration Unit Water		Mixed CO_2 to Stripping Column Reboiler		Mixed CO_2 to Propane Evaporator		Mixed CO_2 to Stripping Column		Stripping Column Overheads		Overheads to Module C		Module C Permeate	
Vapor & Liquid																				
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	94	0.23	0	0.00	122	0.23	0	0.00	122	0.23	122	0.23	122	0.23	122	0.50	122	0.50	28	0.21
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.02	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	27	0.07	0	0.00	27	0.05	0	0.00	25	0.05	25	0.05	25	0.05	2	0.01	2	0.01	0	0.00
CO_2	30,616	74.95	0	0.01	42,224	80.05	0	0.00	41,682	80.13	41,682	80.13	41,682	80.13	14,272	58.09	14,272	58.09	11,609	86.48
H_2O	1,711	4.19	1,523	99.97	188	0.36	152	99.98	13	0.02	13	0.02	13	0.02	0	0.00	0	0.00	0	0.00
N_2	7,864	19.25	0	0.00	9,444	17.91	0	0.00	9,441	18.15	9,441	18.15	9,441	18.15	9,441	38.43	9,441	38.43	1,580	11.77
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	1	0.00	0	0.00	1	0.00	0	0.00	1	0.00	1	0.00	1	0.00	1	0.01	1	0.01	1	0.00
NO_2	1	0.00	0	0.00	1	0.00	0	0.01	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	525	1.28	0	0.00	730	1.38	0	0.00	729	1.40	729	1.40	729	1.40	729	2.97	729	2.97	205	1.53
SO_2	6	0.02	0	0.00	7	0.01	0	0.00	3	0.01	3	0.01	3	0.01	0	0.00	0	0.00	0	0.00
TOTAL	40,846	100.00	1,524	100.00	52,745	100.00	152	100.00	52,015	100.00	52,015	100.00	52,015	100.00	24,567	100.00	24,567	100.00	13,423	100.00
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	3,751	0.23	0	0.00	4,866	0.23	0	0.00	4,861	0.23	4,861	0.23	4,861	0.23	4,860	0.53	4,860	0.53	1,115	0.20
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	41	0.00	40	0.14	2	0.00	0	0.00	2	0.00	2	0.00	2	0.00	0	0.00	0	0.00	0	0.00
Cl_2	1,929	0.12	0	0.00	1,929	0.09	0	0.00	1,764	0.08	1,764	0.08	1,764	0.08	125	0.01	125	0.01	0	0.00
CO_2	1,347,400	83.09	9	0.03	1,858,290	86.16	0	0.00	1,834,400	86.15	1,834,400	86.15	1,834,400	86.15	628,105	68.20	628,105	68.20	510,898	90.77
H_2O	30,829	1.90	27,446	99.82	3,388	0.16	2,734	99.96	227	0.01	227	0.01	227	0.01	4	0.00	4	0.00	4	0.00
N_2	220,293	13.59	0	0.00	264,562	12.27	0	0.00	264,468	12.42	264,468	12.42	264,468	12.42	264,468	28.72	264,468	28.72	44,270	7.86
NH_3	6	0.00	1	0.00	6	0.00	0	0.00	5	0.00	5	0.00	5	0.00	0	0.00	0	0.00	0	0.00
NO	29	0.00	0	0.00	45	0.00	0	0.00	45	0.00	45	0.00	45	0.00	45	0.00	45	0.00	16	0.00
NO_2	45	0.00	0	0.00	44	0.00	1	0.03	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	16,787	1.04	0	0.00	23,349	1.08	0	0.00	23,327	1.10	23,327	1.10	23,327	1.10	23,318	2.53	23,318	2.53	6,563	1.17
SO_2	416	0.03	0	0.00	421	0.02	0	0.00	182	0.01	182	0.01	182	0.01	6	0.00	6	0.00	5	0.00
TOTAL	1,621,527	100.00	27,496	100.00	2,156,902	100.00	2,735	100.00	2,129,281	100.00	2,129,281	100.00	2,129,281	100.00	920,932	100.00	920,932	100.00	562,871	100.00
Solid																				
Solid Components	**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0		0		0	
All Phases																				
All Phases																				
Temperature, °F	95		95		92		95		141		67		-22		-20		35		35	
Pressure, psia	16.2		32.1		438.6		16.8		428.4		428.3		418.3		421.0		416.0		130.5	
Total Flow, lb/h	1,621,527		27,496		2,156,902		2,735		2,129,281		2,129,281		2,129,281		920,932		920,932		562,871	
Enthalpy, Btu/lb (V&L Only)	-3,301.1		-6,846.9		-3,329.0		-6,844.8		-3,307.9		-3,326.0		-3,423.2		-2,654.8		-2,640.6		-3,501.6	
Flow Rate, MMSCFD (V Only)	372				480				474		474		197		224		224		122	
Vapor Frac	1.00		0.00		1.00		0.00		1.00		1.00		0.42		1.00		1.00		1.00	
Density, lb/ft^3 (V&L Only)	0.109		61.446		3.446		61.457		2.980		3.597		9.576		4.040		3.309		1.091	
Average MW (V&L Only)	39.70		18.02		40.89		18.02		40.94		40.94		40.94		37.49		37.49		41.93	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 14.696 psia and 60 °F.
3. Enthalpies are referenced to the constituent elements in their standard states at 77 °F and 14.696 psia.

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EPRI / MTR CO_2 MEMBRANE DESIGN CASE

STREAM	50		51		52		53		54		55		56		57		58	
DESCRIPTION	Module C Recycle Gas to Expander		Module C Recycle Gas to Refrigeration Heat Exchanger 4		Module C Recycle Gas to Refrigeration Heat Exchanger 1		Module C Recycle Gas to Flue Gas Mixing Point		Stripping Column Bottoms		Pumped CO_2 to Refrigeration Heat Exchanger 2		Pumped CO_2 Product		Propane to Condenser		Propane to Refrigeration Heat Exchanger 2	
Vapor & Liquid																		
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	94	0.84	94	0.84	94	0.84	94	0.84	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	37,147	100.00	37,147	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	2	0.02	2	0.02	2	0.02	2	0.02	23	0.08	23	0.08	23	0.08	0	0.00	0	0.00
CO_2	2,663	23.90	2,663	23.90	2,663	23.90	2,663	23.90	27,410	99.86	27,410	99.86	27,410	99.86	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	12	0.04	12	0.04	12	0.04	0	0.00	0	0.00
N_2	7,860	70.54	7,860	70.54	7,860	70.54	7,860	70.54	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	1	0.01	1	0.01	1	0.01	1	0.01	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	524	4.70	524	4.70	524	4.70	524	4.70	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	3	0.01	3	0.01	3	0.01	0	0.00	0	0.00
TOTAL	11,144	100.00	11,144	100.00	11,144	100.00	11,144	100.00	27,448	100.00	27,448	100.00	27,448	100.00	37,147	100.00	37,147	100.00
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	3,745	1.05	3,745	1.05	3,745	1.05	3,745	1.05	1	0.00	1	0.00	1	0.00	0	0.00	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	1,638,040	100.00	1,638,040	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	2	0.00	2	0.00	2	0.00	0	0.00	0	0.00
Cl_2	125	0.03	125	0.03	125	0.03	125	0.03	1,639	0.14	1,639	0.14	1,639	0.14	0	0.00	0	0.00
CO_2	117,208	32.73	117,208	32.73	117,208	32.73	117,208	32.73	1,206,300	99.83	1,206,300	99.83	1,206,300	99.83	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	222	0.02	222	0.02	222	0.02	0	0.00	0	0.00
N_2	220,198	61.50	220,198	61.50	220,198	61.50	220,198	61.50	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	5	0.00	5	0.00	5	0.00	0	0.00	0	0.00
NO	29	0.01	29	0.01	29	0.01	29	0.01	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	16,755	4.68	16,755	4.68	16,755	4.68	16,755	4.68	9	0.00	9	0.00	9	0.00	0	0.00	0	0.00
SO_2	1	0.00	1	0.00	1	0.00	1	0.00	176	0.01	176	0.01	176	0.01	0	0.00	0	0.00
TOTAL	358,062	100.00	358,062	100.00	358,062	100.00	358,062	100.00	1,208,354	100.00	1,208,354	100.00	1,208,354	100.00	1,638,040	100.00	1,638,040	100.00
Solid																		
Solid Components	**lb/h**		**lb/h**				**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0		0	
$CaSO_4 \cdot 2H_2O$	0		0		0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0		0	
All Phases																		
All Phases																		
Temperature, °F	35		-131		5		105		21		45		70		117		95	
Pressure, psia	387.3		31.4		30.4		29.4		422.1		2219.7		2214.7		183.8		176.8	
Total Flow, lb/h	358,062		358,062		358,062		358,062		1,208,354		1,208,354		1,208,354		1,638,040		1,638,040	
Enthalpy, Btu/lb (V&L Only)	-1,273.5		-1,306.0		-1,275.2		-1,252.1		-3,976.9		-3,968.3		-3,955.9		-1,018.6		-1,165.6	
Flow Rate, MMSCFD (V Only)	101		101		101		101		N/A		N/A		N/A		338		N/A	
Vapor Frac	1.00		1.00		1.00		1.00		0.00		0.00		0.00		1.00		0.00	
Density, lb/ft^3 (V&L Only)	2.458		0.290		0.197		0.156		60.138		55.306		48.129		1.626		29.720	
Average MW (V&L Only)	32.13		32.13		32.13		32.13		44.02		44.02		44.02		44.10		44.10	

Notes:

1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 14.696 psia and 60 °F.
3. Enthalpies are referenced to the constituent elements in their standard states at 77 °F and 14.696 psia.

Rev A
11-Jun-15

EPRI / MTR CO_2 MEMBRANE DESIGN CASE

STREAM	59		60		61		62		63		64	
DESCRIPTION	Propane to Refrigeration Heat Exchanger 3		Propane to Refrigeration Heat Exchanger 4		Propane to Letdown Valve		Propane to Evaporator		Propane to Compressor		Propane to Refrigeration Heat Exchanger 1	
Vapor & Liquid												
V&L Mixture Component	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**	**lbmol/h**	**Mole %**
Ar	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
C_3H_8	37,147	100.00	37,147	100.00	37,147	100.00	37,147	100.00	37,147	100.00	37,147	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
N_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
TOTAL	37,147	100.00	37,147	100.00	37,147	100.00	37146.69	100.00	37146.69	100.00	37146.69	100.00
V&L Mixture Component	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**	**lb/h**	**Mass %**
Ar	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
C_3H_8	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
N_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
TOTAL	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00	1,638,040	100.00
Solid												
Solid Components	**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**		**lb/h**	
Coal	0		0		0		0		0		0	
Ash	0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0	
All Phases												
All Phases												
Temperature, °F	82		70		59		-31		-31		128	
Pressure, psia	175.5		174.3		173.0		20.0		19.9		185.0	
Total Flow, lb/h	1,638,040		1,638,040		1,638,040		1,638,040		1,638,040		1,638,040	
Enthalpy, Btu/lb (V&L Only)	-1,174.7		-1,182.7		-1,189.4		-1,189.4		-1,063.2		-1,013.6	
Flow Rate, MMSCFD (V Only)	N/A		N/A		N/A		100		338		338	
Vapor Frac	0.00		0.00		0.00		0.29		1.00		1.00	
Density, lb/ft^3 (V&L Only)	30.472		31.125		31.670		0.670		0.199		1.586	
Average MW (V&L Only)	44.10		44.10		44.10		44.10		44.10		44.10	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 14.696 psia and 60 °F.
3. Enthalpies are referenced to the constituent elements in their standard states at 77 °F and 14.696 psia.

Rev A
11-Jun-15

WorleyParsons
resources & energy

EPRI / MTR CO_2 MEMBRANE DESIGN CASE

STREAM	1		2		3		4		5		6		7	
DESCRIPTION	Ambient Air to Primary Air Fan		Primary Air		Primary Air Leakage to Flue Gas		Primary Air to Coal Milling		Coal Feed		Infiltration Air		Bottom Ash	
Vapor & Liquid														
V&L Mixture Component	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**
Ar	190	0.92	190	0.92	26	0.92	164	0.92	0	N/A	16	0.92	0	N/A
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
CO_2	7	0.03	7	0.03	1	0.03	6	0.03	0	N/A	1	0.03	0	N/A
H_2O	203	0.99	203	0.99	28	0.99	175	0.99	0	N/A	17	0.99	0	N/A
N_2	15,887	77.32	15,887	77.32	2,186	77.32	13,700	77.32	0	N/A	1,348	77.32	0	N/A
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
O_2	4,262	20.74	4,262	20.74	587	20.74	3,675	20.74	0	N/A	362	20.74	0	N/A
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
TOTAL	20,548	100.00	20,548	100	2,828	100	17,720	100	0	N/A	1,744	100	0	N/A
V&L Mixture Component	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**
Ar	7,593	1.28	7,593	1.28	1,045	1.28	6,548	1.28	0	N/A	644	1.28	0	N/A
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
CO_2	297	0.05	297	0.05	41	0.05	256	0.05	0	N/A	25	0.05	0	N/A
H_2O	3,654	0.62	3,654	0.62	503	0.62	3,151	0.62	0	N/A	310	0.62	0	N/A
N_2	445,037	75.06	445,037	75.06	61,249	75.06	383,788	75.06	0	N/A	37,773	75.06	0	N/A
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
O_2	136,365	23.00	136,365	23.00	18,767	23.00	117,597	23.00	0	N/A	11,574	23.00	0	N/A
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	N/A
TOTAL	592,945	100	592,945	100	81,605	100	511,339	100	0	N/A	50,327	100	0	N/A
Solid														
Solid Components	**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**	
Coal	0		0		0		0		256,711		0		0	
Ash	0		0		0		0		0		0		4,980	
$CaCO_3$	0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0	
$CaSO_4 \cdot 2H_2O$	0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0	
TOTAL	0		0		0		0		256,711		0		4,980	
All Phases														
All Phases														
Temperature, °C	15		25		25		243		15		15		375	
Pressure, MPa	0.101		0.111		0.111		0.111		0.101		0.101		0.101	
Total Flow, kg/h	592,945		592,945		81,605		511,339		256,711		50,327		4,980	
Enthalpy, kJ/kg (V&L Only)	-97.6		-87.3		-87.3		135.1		N/A		-97.6		N/A	
Flow Rate, Nm^3/min (V Only)	8,113		8,113		1,117		6,996		N/A		689		N/A	
Vapor Frac	1.00		1.00		1.00		1.00		N/A		1.00		N/A	
Density, kg/m^3 (V&L Only)	1.22		1.29		1.29		0.75		N/A		1.22		N/A	
Average MW (V&L Only)	28.86		28.86		28.86		28.86		N/A		28.86		N/A	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 1.01 bar and 15.6 °C.
3. Enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.

Rev A
11-Jun-15

EPRI / MTR CO_2 MEMBRANE
DESIGN CASE

STREAM	8		9		10		11		12		13		14		15		16		17		18		19	
DESCRIPTION	**Ambient Air to Secondary Air Booster Fan**		**Secondary Air to Module B**		**Vitiated Secondary Air to Forced Draft Fan**		**Secondary Air to Air Heater**		**Secondary Air Leakage to Flue Gas**		**Secondary Air to Boiler**		**Flue Gas to Air Heater**		**Flue Gas to Fabric Filter**		**Fly Ash**		**Flue Gas to ID Fans**		**ID Fan Discharge**		**TEG Dehydration Unit Vent Recycle**	
Vapor & Liquid																								
V&L Mixture Component	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**
Ar	642	0.92	642	0.92	648	0.86	648	0.86	19	0.86	629	0.86	809	0.83	854	0.83	0	N/A	854	0.83	854	0.83	0	0.02
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.05
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	10	0.01	10	0.01	0	N/A	10	0.01	10	0.01	1	0.40
CO_2	23	0.03	23	0.03	5,694	7.56	5,694	7.56	169	7.56	5,526	7.56	19,157	19.61	19,327	18.81	0	N/A	19,327	18.81	19,327	18.81	246	93.91
H_2O	685	0.99	685	0.99	895	1.19	895	1.19	26	1.19	868	1.19	8,418	8.62	8,472	8.25	0	N/A	8,472	8.25	8,472	8.25	11	4.10
N_2	53,641	77.32	53,641	77.32	54,190	71.99	54,190	71.99	1,605	71.99	52,585	71.99	67,760	69.37	71,552	69.65	0	N/A	71,552	69.65	71,552	69.65	2	0.58
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	4	0.00	4	0.00	0	N/A	4	0.00	4	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.16
O_2	14,389	20.74	14,389	20.74	13,850	18.40	13,850	18.40	410	18.40	13,440	18.40	1,314	1.34	2,310	2.25	0	N/A	2,310	2.25	2,310	2.25	0	0.12
SO_2	0	0.00	0	0.00	1	0.00	1	0.00	0	0.00	1	0.00	202	0.21	202	0.20	0	N/A	202	0.20	202	0.20	2	0.65
TOTAL	69,380	100.00	69,380	100.00	75,278	100.00	75,278	100.00	2,230	100.00	73,049	100.00	97,675	100.00	102,733	100.00	0	N/A	102,733	100.00	102,733	100.00	262	100.00
V&L Mixture Component	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**
Ar	25,637	1.28	25,637	1.28	25,896	1.15	25,896	1.15	767	1.15	25,129	1.15	32,321	1.08	34,133	1.09	0	N/A	34,133	1.09	34,133	1.09	2	0.02
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	N/A	0	0.00	0	0.00	19	0.17
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	744	0.02	744	0.02	0	N/A	744	0.02	744	0.02	75	0.66
CO_2	1,002	0.05	1,002	0.05	250,599	11.12	250,599	11.12	7,422	11.12	243,176	11.12	843,103	28.28	850,566	27.18	0	N/A	850,566	27.18	850,566	27.18	10,837	95.84
H_2O	12,336	0.62	12,336	0.62	16,115	0.71	16,115	0.71	477	0.71	15,638	0.71	151,651	5.09	152,631	4.88	0	N/A	152,631	4.88	152,631	4.88	194	1.71
N_2	1,502,680	75.06	1,502,680	75.06	1,518,050	67.35	1,518,050	67.35	44,962	67.35	1,473,090	67.35	1,898,200	63.67	2,004,410	64.05	0	N/A	2,004,410	64.05	2,004,410	64.05	43	0.38
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	3	0.00	3	0.00	0	N/A	3	0.00	3	0.00	0	0.00
NO	0	0.00	0	0.00	12	0.00	12	0.00	0	0.00	12	0.00	132	0.00	132	0.00	0	N/A	132	0.00	132	0.00	0	0.00
NO_2	0	0.00	0	0.00	8	0.00	8	0.00	0	0.00	8	0.00	12	0.00	12	0.00	0	N/A	12	0.00	12	0.00	20	0.17
O_2	460,439	23.00	460,439	23.00	443,181	19.66	443,181	19.66	13,126	19.66	430,055	19.66	42,032	1.41	73,926	2.36	0	N/A	73,926	2.36	73,926	2.36	10	0.09
SO_2	0	0.00	0	0.00	62	0.00	62	0.00	2	0.00	60	0.00	12,933	0.43	12,935	0.41	0	N/A	12,935	0.41	12,935	0.41	108	0.96
TOTAL	2,002,094	100.00	2,002,094	100.00	2,253,923	100.00	2,253,923	100.00	66,757	100.00	2,187,167	100.00	2,981,130	100.00	3,129,491	100.00	0	N/A	3,129,491	100.00	3,129,491	100.00	11,308	100.00
Solid																								
Solid Components	**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**	
Coal	0		0		0		0		0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		19,921		19,921		19,921		0		0		0	
$CaCO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		19,921		19,921		19,921		0		0		0	
All Phases																								
All Phases																								
Temperature, °C	15		24		35		35		35		221		378		211		211		211		223		35	
Pressure, MPa	0.101		0.110		0.105		0.105		0.105		0.103		0.100		0.099		0.098		0.098		0.105		0.116	
Total Flow, kg/h	2,002,094		2,002,094		2,253,923		2,253,923		66,757		2,187,167		3,001,052		3,149,413		19,921		3,129,491		3,129,491		11,308	
Enthalpy, kJ/kg (V&L Only)	-97.6		-88.4		-1,081.1		-1,080.6		-1,080.6		-892.0		-2,852.3		-2,909.5		N/A		-2,909.5		-2,896.2		-8,842.8	
Flow Rate, Nm^3/min (V Only)	27,393		27,393		29,722		29,722		880		28,842		38,565		40,562		N/A		40,562		40,562		104	
Vapor Frac	1.00		1.00		1.00		1.00		1.00		1.00		1.00		1.00		N/A		1.00		1.00		1.00	
Density, kg/m^3 (V&L Only)	1.22		1.29		1.23		1.23		1.23		0.75		0.56		0.75		N/A		0.74		0.78		1.96	
Average MW (V&L Only)	28.86		28.86		29.94		29.94		29.94		29.94		30.52		30.46		N/A		30.46		30.46		43.13	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 1.01 bar and 15.6 °C.
3. Enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.

Rev A
11-Jun-15

EPRI / MTR CO_2 MEMBRANE
DESIGN CASE

STREAM	20		21		22		23		24		25		26		27		28		29		30		31	
DESCRIPTION	Flue Gas to Flue Gas Glycol Cooler		Flue Gas to FGD		FGD Limestone Slurry		FGD Make-Up Water		FGD Oxidation Air		Gypsum		Desulfurized Flue Gas		Flue Gas from Compressor		Flue Gas to Direct-Contact Cooler		Direct Contact Cooler Blowdown Water		Flue Gas to Module A		Flue Gas to Module B	
Vapor & Liquid																								
V&L Mixture Component	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %
Ar	854	0.83	854	0.83	0	0.00	0	0.00	9	0.92	0	0.00	864	0.77	864	0.77	906	0.77	0	0.00	906	0.88	864	1.04
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	12	0.01	12	0.01	0	0.00	0	0.00	0	0.00	0	0.00	12	0.01	12	0.01	12	0.01	0	0.00	12	0.01	0	0.00
CO_2	19,573	19.00	19,573	19.00	0	0.00	0	0.00	0	0.03	0	0.01	19,773	17.61	19,773	17.61	20,981	17.88	0	0.00	20,980	20.38	7,093	8.51
H_2O	8,483	8.24	8,483	8.24	3,424	100.00	13,871	100.00	10	0.99	268	99.99	16,886	15.04	16,886	15.04	16,886	14.39	14,420	100.00	2,466	2.40	653	0.78
N_2	71,553	69.47	71,553	69.47	0	0.00	0	0.00	781	77.32	0	0.00	72,334	64.41	72,334	64.41	75,900	64.68	0	0.00	75,900	73.74	72,333	86.76
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	4	0.00	4	0.00	0	0.00	0	0.00	0	0.00	0	0.00	4	0.00	4	0.00	5	0.00	0	0.00	5	0.00	4	0.01
NO_2	1	0.00	1	0.00	0	0.00	0	0.00	0	0.00	0	0.00	1	0.00	1	0.00	1	0.00	0	0.00	1	0.00	0	0.00
O_2	2,311	2.24	2,311	2.24	0	0.00	0	0.00	210	20.74	0	0.00	2,420	2.16	2,420	2.16	2,658	2.26	0	0.00	2,658	2.58	2,420	2.90
SO_2	204	0.20	204	0.20	0	0.00	0	0.00	0	0.00	0	0.00	4	0.00	4	0.00	4	0.00	0	0.00	4	0.00	1	0.00
TOTAL	102,995	100.00	102,995	100.00	3,424	100.00	13,871	100.00	1,010	100.00	267.65	100.00	112,298	100.00	112,298	100.00	117,353	100.00	14,420	100.00	102,933	100.00	83,369	100.00
V&L Mixture Component	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %
Ar	34,135	1.09	34,135	1.09	0	0.00	0	0.00	373	1.28	0	0.00	34,508	1.04	34,508	1.04	36,207	1.04	0	0.00	36,207	1.13	34,506	1.40
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	19	0.00	19	0.00	0	0.00	0	0.00	0	0.00	0	0.00	19	0.00	19	0.00	19	0.00	0	0.00	19	0.00	0	0.00
Cl_2	819	0.03	819	0.03	0	0.00	0	0.00	0	0.00	0	0.00	819	0.02	819	0.02	875	0.03	0	0.00	875	0.03	0	0.00
CO_2	861,403	27.43	861,403	27.43	0	0.00	0	0.00	15	0.05	1	0.03	870,200	26.26	870,200	26.26	923,349	26.56	4	0.00	923,345	28.71	312,173	12.68
H_2O	152,825	4.87	152,825	4.87	61,688	100.00	249,884	100.00	180	0.62	4,821	99.97	304,205	9.18	304,205	9.18	304,205	8.75	259,778	100.00	44,427	1.38	11,770	0.48
N_2	2,004,450	63.82	2,004,450	63.82	0	0.00	0	0.00	21,882	75.06	0	0.00	2,026,330	61.15	2,026,330	61.15	2,126,220	61.16	0	0.00	2,126,220	66.10	2,026,300	82.29
NH_3	3	0.00	3	0.00	0	0.00	0	0.00	0	0.00	0	0.00	3	0.00	3	0.00	3	0.00	1	0.00	3	0.00	0	0.00
NO	132	0.00	132	0.00	0	0.00	0	0.00	0	0.00	0	0.00	132	0.00	132	0.00	145	0.00	0	0.00	145	0.00	132	0.01
NO_2	31	0.00	31	0.00	0	0.00	0	0.00	0	0.00	0	0.00	31	0.00	31	0.00	31	0.00	1	0.00	31	0.00	10	0.00
O_2	73,936	2.35	73,936	2.35	0	0.00	0	0.00	6,705	23.00	0	0.00	77,448	2.34	77,448	2.34	85,046	2.45	0	0.00	85,046	2.64	77,431	3.14
SO_2	13,043	0.42	13,043	0.42	0	0.00	0	0.00	0	0.00	0	0.00	256	0.01	256	0.01	257	0.01	0	0.00	257	0.01	68	0.00
TOTAL	3,140,797	100.00	3,140,797	100.00	61,688	100.00	249,884	100.00	29,154	100.00	4,823	100.00	3,313,951	100.00	3,313,951	100.00	3,476,358	100.00	259,783	100.00	3,216,575	100.00	2,462,391	100.00
Solid																								
Solid Components	kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h	
Coal	0		0		0		0		0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		25,314		0		0		5,337		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		34,364		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0		0		0		0		0	
SiO_2	0		0		799		0		0		799		0		0		0		0		0		0	
Fe_2O_3	0		0		533		0		0		533		0		0		0		0		0		0	
TOTAL	0		0		26,646		0		0		41,034		0		0		0		0		0		0	
All Phases																								
All Phases																								
Temperature, °C	223		167		15		15		15		57		57		134		130		35		35		35	
Pressure, MPa	0.105		0.105		0.138		0.345		0.101		0.102		0.102		0.202		0.202		0.200		0.200		0.195	
Total Flow, kg/h	3,140,797		3,140,797		88,334		249,884		29,154		45,856		3,313,951		3,313,951		3,476,358		259,783		3,216,575		2,462,391	
Enthalpy, kJ/kg (V&L Only)	-2,917.6		-2,977.6		-16,017.3		-16,017.1		-97.6		-15,825.4		-3,547.0		-3,464.1		-3,438.3		-15,926.6		-2,743.3		-1,188.2	
Flow Rate, Nm^3/min (V Only)	40,666		40,666		N/A		N/A		399		N/A		44,339		44,339		46,335		N/A		40,641		32,917	
Vapor Frac	1.00		1.00		0.00		0.00		1.00		0.00		1.00		1.00		1.00		0.00		1.00		1.00	
Density, kg/m^3 (V&L Only)	0.78		0.88		1003.55		1003.55		1.22		829.31		1.09		1.77		1.79		984.26		2.45		2.25	
Average MW (V&L Only)	30.49		30.49		18.02		18.02		28.86		18.02		29.51		29.51		29.62		18.02		31.25		29.54	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 1.01 bar and 15.6 °C.
3. Enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.

Rev A
11-Jun-15

WorleyParsons
resources & energy

EPRI / MTR CO_2 MEMBRANE
DESIGN CASE

STREAM	32		33		34		35		36		37		38		39	
DESCRIPTION	Flue Gas to Flue Gas Glycol Heater		Flue Gas to Expander		Flue Gas to Stack		Module A Permeate		Condensed Water from Vacuum Compressor		Raw CO_2 to Cooler		Raw CO_2 to Knock Out Drum		Condensed Water from Knock Out Drum	
Vapor & Liquid																
V&L Mixture Component	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**
Ar	857	1.11	857	1.11	857	1.11	43	0.22	0	0.00	43	0.23	43	0.23	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	12	0.06	0	0.00	12	0.07	12	0.07	0	0.00
CO_2	1,422	1.84	1,422	1.84	1,422	1.84	13,887	70.98	0	0.00	13,887	73.79	13,887	73.79	0	0.00
H_2O	444	0.57	444	0.57	444	0.57	1,813	9.27	745	100.00	1,068	5.67	1,068	5.67	291	100.00
N_2	71,784	92.66	71,784	92.66	71,784	92.66	3,567	18.23	0	0.00	3,567	18.95	3,567	18.95	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	4	0.01	4	0.01	4	0.01	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	2,959	3.82	2,959	3.82	2,959	3.82	238	1.22	0	0.00	238	1.26	238	1.26	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	3	0.02	0	0.00	3	0.02	3	0.02	0	0.00
TOTAL	77,470	100.00	77,470	100.00	77,470	100.00	19,564	100.00	745	100.00	18,819	100.00	18,819	100.00	291	100.00
V&L Mixture Component	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**
Ar	34,247	1.55	34,247	1.55	34,247	1.55	1,702	0.23	0	0.00	1,702	0.23	1,702	0.23	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	19	0.00	0	0.00	19	0.00	19	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	875	0.12	0	0.00	875	0.12	875	0.12	0	0.00
CO_2	62,576	2.83	62,576	2.83	62,576	2.83	611,172	81.04	0	0.00	611,172	82.51	611,172	82.51	0	0.00
H_2O	7,992	0.36	7,992	0.36	7,992	0.36	32,657	4.33	13,425	100.00	19,231	2.60	19,231	2.60	5,248	100.00
N_2	2,010,920	90.97	2,010,920	90.97	2,010,920	90.97	99,923	13.25	0	0.00	99,923	13.49	99,923	13.49	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	3	0.00	0	0.00	3	0.00	3	0.00	0	0.00
NO	120	0.01	120	0.01	120	0.01	13	0.00	0	0.00	13	0.00	13	0.00	0	0.00
NO_2	2	0.00	2	0.00	2	0.00	20	0.00	0	0.00	20	0.00	20	0.00	0	0.00
O_2	94,689	4.28	94,689	4.28	94,689	4.28	7,614	1.01	0	0.00	7,614	1.03	7,614	1.03	0	0.00
SO_2	6	0.00	6	0.00	6	0.00	189	0.03	0	0.00	189	0.03	189	0.03	0	0.00
TOTAL	2,210,552	100.00	2,210,552	100.00	2,210,552	100.00	754,187	100.00	13,426	100.00	740,761	100.00	740,761	100.00	5,248	100.00
Solid																
Solid Components	**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**	
Coal	0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0	
All Phases																
All Phases																
Temperature, °C	24		107		53		35		35		66		35		35	
Pressure, MPa	0.190		0.190		0.101		0.020		0.020		0.117		0.114		0.112	
Total Flow, kg/h	2,210,552		2,210,552		2,210,552		754,187		13,426		740,761		740,761		5,248	
Enthalpy, kJ/kg (V&L Only)	-302.7		-217.4		-272.7		-7,819.3		-15,926.6		-7,690.4		-7,736.9		-15,927.7	
Flow Rate, Nm^3/min (V Only)	30,588		30,588		30,588		7,724		N/A		7,430		7,314		N/A	
Vapor Frac	1.00		1.00		1.00		1.00		0.00		1.00		0.98		0.00	
Density, kg/m^3 (V&L Only)	2.19		1.71		1.07		0.30		926.71		1.64		1.79		984.52	
Average MW (V&L Only)	28.53		28.53		28.53		38.55		18.02		39.36		39.36		18.02	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 1.01 bar and 15.6 °C.
3. Enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.

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WorleyParsons
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EPRI / MTR CO_2 MEMBRANE DESIGN CASE

STREAM	40		41		42		43		44		45		46		47		48		49	
DESCRIPTION	Raw CO_2 to Multi-Stage Compressor		Condensed Water from Compressor		Raw CO_2 to TEG Unit		TEG Dehydration Unit Water		Mixed CO_2 to Stripping Column Reboiler		Mixed CO_2 to Propane Evaporator		Mixed CO_2 to Stripping Column		Stripping Column Overheads		Overheads to Module C		Module C Permeate	
Vapor & Liquid																				
V&L Mixture Component	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %
Ar	43	0.23	0	0.00	55	0.23	0	0.00	55	0.23	55	0.23	55	0.23	55	0.50	55	0.50	13	0.21
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.02	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	12	0.07	0	0.00	12	0.05	0	0.00	11	0.05	11	0.05	11	0.05	1	0.01	1	0.01	0	0.00
CO_2	13,887	74.95	0	0.01	19,153	80.05	0	0.00	18,906	80.13	18,906	80.13	18,906	80.13	6,474	58.09	6,474	58.09	5,266	86.48
H_2O	776	4.19	691	99.97	85	0.36	69	99.98	6	0.02	6	0.02	6	0.02	0	0.00	0	0.00	0	0.00
N_2	3,567	19.25	0	0.00	4,284	17.91	0	0.00	4,282	18.15	4,282	18.15	4,282	18.15	4,282	38.43	4,282	38.43	717	11.77
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	1	0.00	0	0.00	1	0.00	1	0.00	1	0.00	1	0.01	1	0.01	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.01	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	238	1.28	0	0.00	331	1.38	0	0.00	331	1.40	331	1.40	331	1.40	331	2.97	331	2.97	93	1.53
SO_2	3	0.02	0	0.00	3	0.01	0	0.00	1	0.01	1	0.01	1	0.01	0	0.00	0	0.00	0	0.00
TOTAL	18,527	100.00	691	100.00	23,925	100.00	69	100.00	23,594	100.00	23,594	100.00	23,594	100.00	11,143	100.00	11,143	100.00	6,089	100.00
V&L Mixture Component	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %
Ar	1,702	0.23	0	0.00	2,207	0.23	0	0.00	2,205	0.23	2,205	0.23	2,205	0.23	2,205	0.53	2,205	0.53	506	0.20
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
$C_6H_{14}O_4$ (TEG)	19	0.00	18	0.14	1	0.00	0	0.00	1	0.00	1	0.00	1	0.00	0	0.00	0	0.00	0	0.00
Cl_2	875	0.12	0	0.00	875	0.09	0	0.00	800	0.08	800	0.08	800	0.08	57	0.01	57	0.01	0	0.00
CO_2	611,171	83.09	4	0.03	842,907	86.16	0	0.00	832,070	86.15	832,070	86.15	832,070	86.15	284,904	68.20	284,904	68.20	231,739	90.77
H_2O	13,984	1.90	12,449	99.82	1,537	0.16	1,240	99.96	103	0.01	103	0.01	103	0.01	2	0.00	2	0.00	2	0.00
N_2	99,923	13.59	0	0.00	120,004	12.27	0	0.00	119,961	12.42	119,961	12.42	119,961	12.42	119,961	28.72	119,961	28.72	20,080	7.86
NH_3	3	0.00	0	0.00	3	0.00	0	0.00	2	0.00	2	0.00	2	0.00	0	0.00	0	0.00	0	0.00
NO	13	0.00	0	0.00	20	0.00	0	0.00	20	0.00	20	0.00	20	0.00	20	0.00	20	0.00	7	0.00
NO_2	20	0.00	0	0.00	20	0.00	0	0.03	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	7,614	1.04	0	0.00	10,591	1.08	0	0.00	10,581	1.10	10,581	1.10	10,581	1.10	10,577	2.53	10,577	2.53	2,977	1.17
SO_2	189	0.03	0	0.00	191	0.02	0	0.00	82	0.01	82	0.01	82	0.01	3	0.00	3	0.00	2	0.00
TOTAL	735,513	100.00	12,472	100.00	978,356	100.00	1,241	100.00	965,826	100.00	965,826	100.00	965,826	100.00	417,728	100.00	417,728	100.00	255,313	100.00
Solid																				
Solid Components	kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h		kg/h	
Coal	0		0		0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0		0		0	
$CaSO_4 \cdot 2H_2O$	0		0		0		0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0		0		0	
All Phases																				
All Phases																				
Temperature, °C	35		35		33		35		60		20		-30		-29		2		2	
Pressure, MPa	0.112		0.221		3.024		0.116		2.954		2.953		2.884		2.903		2.868		0.900	
Total Flow, kg/h	735,513		12,472		978,356		1,241		965,826		965,826		965,826		417,728		417,728		255,313	
Enthalpy, kJ/kg (V&L Only)	-7,678.5		-15,926.0		-7,743.2		-15,921.1		-7,694.2		-7,736.3		-7,962.3		-6,175.0		-6,141.9		-8,144.7	
Flow Rate, Nm^3/min (V Only)	7,315				9,446				9,316		9,316		3,875		4,400		4,400		2,404	
Vapor Frac	1.00		0.00		1.00		0.00		1.00		1.00		0.42		1.00		1.00		1.00	
Density, kg/m^3 (V&L Only)	1.74		984.27		55.19		984.45		47.74		57.62		153.40		64.72		53.01		17.47	
Average MW (V&L Only)	39.70		18.02		40.89		18.02		40.94		40.94		40.94		37.49		37.49		41.93	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 1.01 bar and 15.6 °C.
3. Enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.

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EPRI / MTR CO_2 MEMBRANE DESIGN CASE

STREAM	50		51		52		53		54		55		56		57		58	
DESCRIPTION	Module C Recycle Gas to Expander		Module C Recycle Gas to Refrigeration Heat Exchanger 4		Module C Recycle Gas to Refrigeration Heat Exchanger 1		Module C Recycle Gas to Flue Gas Mixing Point		Stripping Column Bottoms		Pumped CO_2 to Refrigeration Heat Exchanger 2		Pumped CO_2 Product		Propane to Condenser		Propane to Refrigeration Heat Exchanger 2	
Vapor & Liquid																		
V&L Mixture Component	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %	kgmol/h	Mole %
Ar	43	0.84	43	0.84	43	0.84	43	0.84	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	16,849	100.00	16,849	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	1	0.02	1	0.02	1	0.02	1	0.02	10	0.08	10	0.08	10	0.08	0	0.00	0	0.00
CO_2	1,208	23.90	1,208	23.90	1,208	23.90	1,208	23.90	12,433	99.86	12,433	99.86	12,433	99.86	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	6	0.04	6	0.04	6	0.04	0	0.00	0	0.00
N_2	3,565	70.54	3,565	70.54	3,565	70.54	3,565	70.54	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.01	0	0.01	0	0.01	0	0.01	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	238	4.70	238	4.70	238	4.70	238	4.70	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	1	0.01	1	0.01	1	0.01	0	0.00	0	0.00
TOTAL	5,055	100.00	5,055	100.00	5,055	100.00	5,055	100.00	12,450	100.00	12,450	100.00	12,450	100.00	16,849	100.00	16,849	100.00
V&L Mixture Component	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %	kg/h	Mass %
Ar	1,699	1.05	1,699	1.05	1,699	1.05	1,699	1.05	1	0.00	1	0.00	1	0.00	0	0.00	0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	743,002	100.00	743,002	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	1	0.00	1	0.00	1	0.00	0	0.00	0	0.00
Cl_2	57	0.03	57	0.03	57	0.03	57	0.03	744	0.14	744	0.14	744	0.14	0	0.00	0	0.00
CO_2	53,164	32.73	53,164	32.73	53,164	32.73	53,164	32.73	547,166	99.83	547,166	99.83	547,166	99.83	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	101	0.02	101	0.02	101	0.02	0	0.00	0	0.00
N_2	99,880	61.50	99,880	61.50	99,880	61.50	99,880	61.50	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	2	0.00	2	0.00	2	0.00	0	0.00	0	0.00
NO	13	0.01	13	0.01	13	0.01	13	0.01	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	7,600	4.68	7,600	4.68	7,600	4.68	7,600	4.68	4	0.00	4	0.00	4	0.00	0	0.00	0	0.00
SO_2	1	0.00	1	0.00	1	0.00	1	0.00	80	0.01	80	0.01	80	0.01	0	0.00	0	0.00
TOTAL	162,414	100.00	162,414	100.00	162,414	100.00	162,414	100.00	548,098	100.00	548,098	100.00	548,098	100.00	743,002	100.00	743,002	100.00
Solid																		
Solid Components	kg/h		kg/h				kg/h		kg/h		kg/h		kg/h		kg/h		kg/h	
Coal	0		0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0		0		0		0	
$CaSO_4{\cdot}2H_2O$	0		0		0		0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0		0	
All Phases																		
All Phases																		
Temperature, °C	2		-91		-15		41		-6		7		21		47		35	
Pressure, MPa	2.670		0.216		0.209		0.202		2.910		15.304		15.270		1.267		1.219	
Total Flow, kg/h	162,414		162,414		162,414		162,414		548,098		548,098		548,098		743,002		743,002	
Enthalpy, kJ/kg (V&L Only)	-2,962.2		-3,037.7		-2,966.1		-2,912.4		-9,250.2		-9,230.2		-9,201.4		-2,369.3		-2,711.1	
Flow Rate, Nm^3/min (V Only)	1,996		1,996		1,996		1,996		N/A		N/A		N/A		6,653		N/A	
Vapor Frac	1.00		1.00		1.00		1.00		0.00		0.00		0.00		1.00		0.00	
Density, kg/m^3 (V&L Only)	39.37		4.65		3.15		2.50		963.32		885.92		770.96		26.05		476.07	
Average MW (V&L Only)	32.13		32.13		32.13		32.13		44.02		44.02		44.02		44.10		44.10	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 1.01 bar and 15.6 °C.
3. Enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.

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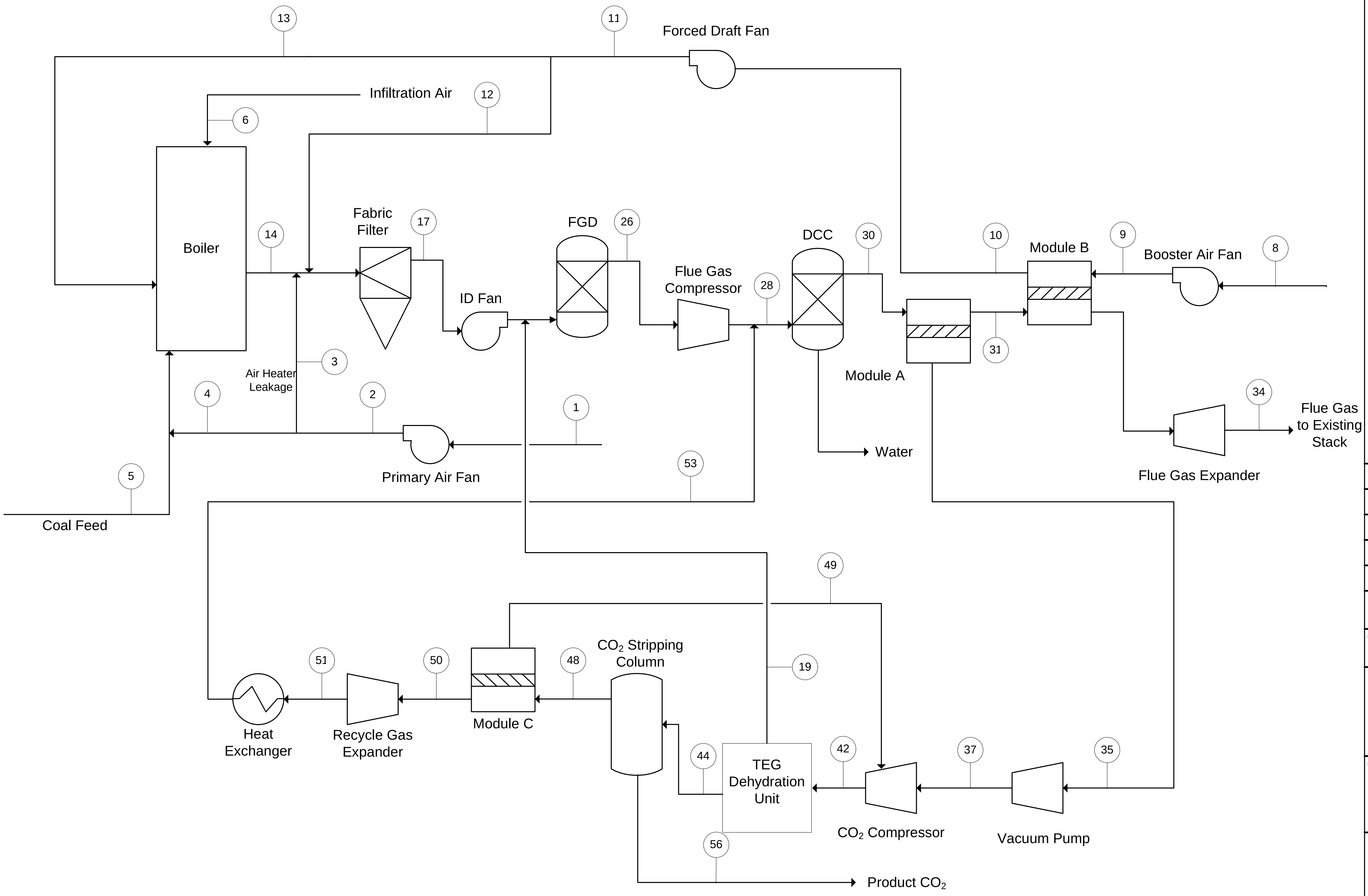
EPRI / MTR CO_2 MEMBRANE DESIGN CASE

STREAM	59		60		61		62		63		64	
DESCRIPTION	Propane to Refrigeration Heat Exchanger 3		Propane to Refrigeration Heat Exchanger 4		Propane to Letdown Valve		Propane to Evaporator		Propane to Compressor		Propane to Refrigeration Heat Exchanger 1	
Vapor & Liquid												
V&L Mixture Component	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**	**kgmol/h**	**Mole %**
Ar	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
C_3H_8	16,849	100.00	16,849	100.00	16,849	100.00	16,849	100.00	16,849	100.00	16,849	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
N_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
TOTAL	16,849	100.00	16,849	100.00	16,849	100.00	16849.46	100.00	16849.46	100.00	16849.46	100.00
V&L Mixture Component	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**	**kg/h**	**Mass %**
Ar	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
C_3H_8	743,002	100.00	743,002	100.00	743,002	100.00	743,002	100.00	743,002	100.00	743,002	100.00
$C_6H_{14}O_4$ (TEG)	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
Cl_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
H_2O	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
N_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NH_3	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
TOTAL	743,002	100.00	743,002	100.00	743,002	100.00	743,002	100.00	743,002	100.00	743,002	100.00
Solid												
Solid Components	**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**		**kg/h**	
Coal	0		0		0		0		0		0	
Ash	0		0		0		0		0		0	
$CaCO_3$	0		0		0		0		0		0	
$CaSO_3$	0		0		0		0		0		0	
$CaSO_4 \cdot 2H_2O$	0		0		0		0		0		0	
$MgCO_3$	0		0		0		0		0		0	
$MgSO_3$	0		0		0		0		0		0	
$MgSO_4$	0		0		0		0		0		0	
SiO_2	0		0		0		0		0		0	
Fe_2O_3	0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0	
All Phases												
All Phases												
Temperature, °C	28		21		15		-35		-35		53	
Pressure, MPa	1.210		1.201		1.193		0.138		0.137		1.276	
Total Flow, kg/h	743,002		743,002		743,002		743,002		743,002		743,002	
Enthalpy, kJ/kg (V&L Only)	-2,732.4		-2,751.0		-2,766.6		-2,766.6		-2,472.9		-2,357.6	
Flow Rate, Nm^3/min (V Only)	N/A		N/A		N/A		1,960		6,648		6,653	
Vapor Frac	0.00		0.00		0.00		0.29		1.00		1.00	
Density, kg/m^3 (V&L Only)	488.12		498.57		507.31		10.74		3.19		25.40	
Average MW (V&L Only)	44.10		44.10		44.10		44.10		44.10		44.10	

Notes:
1. Results based on MTR membrane performance received on 13-May-2015.
2. Standard volumetric flow rates are referenced to 1.01 bar and 15.6 °C.
3. Enthalpies are referenced to the constituent elements in their standard states at 25 °C and 0.101325 MPa.

Rev A
11-Jun-15

Appendix 2: Block Flow Diagram and Process Flow Diagrams



NOTES:

A	06/10/2015	INITIAL ISSUE
REV	DATE	DESCRIPTION

ENGINEER/TECH SPECIALIST: E. MACE

DATE: 06/10/2015

CHECKED BY:

PROJECT MANAGER: D. STAUFFER

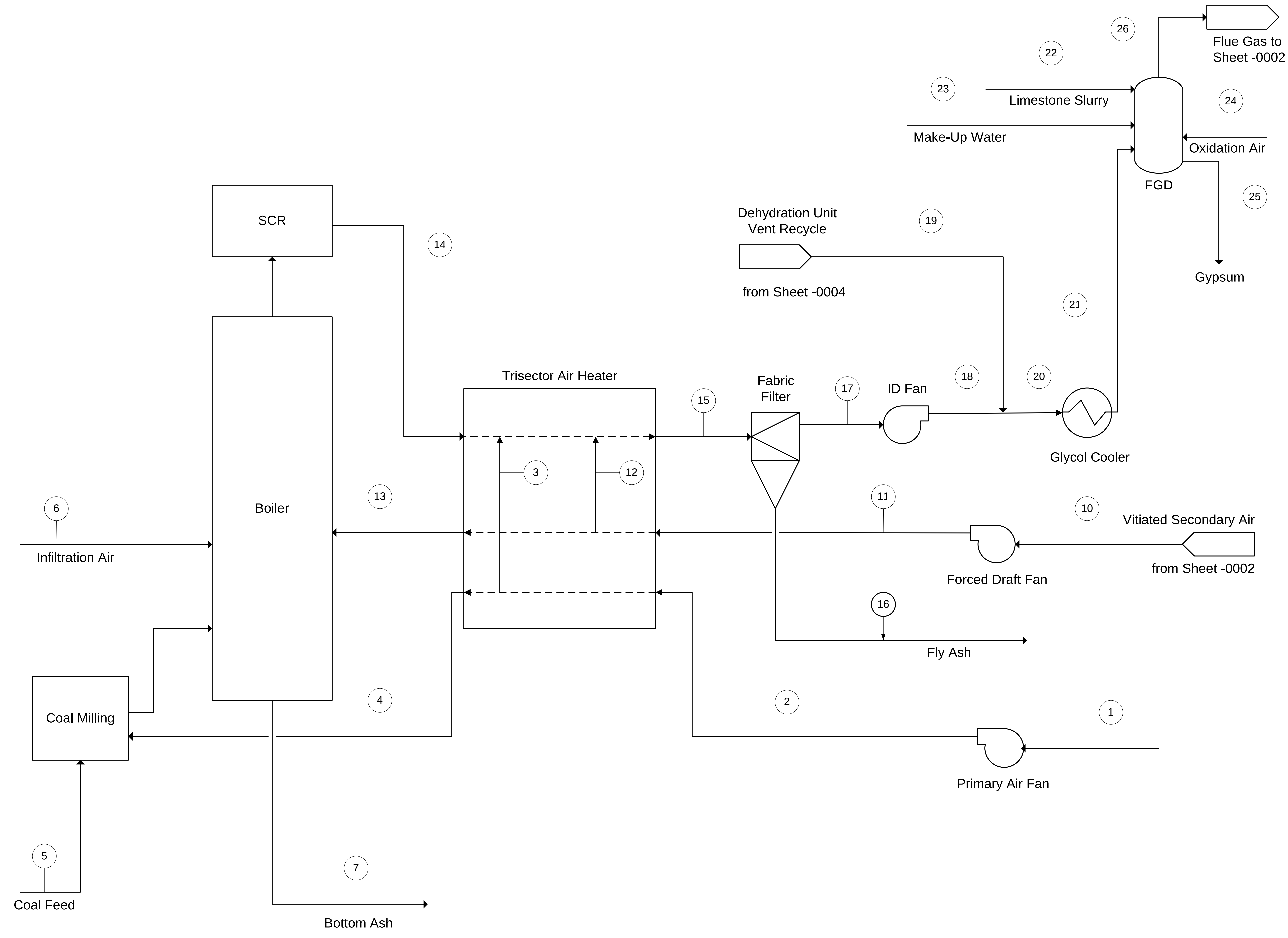
WorleyParsons
resources & energy

CLIENT/PROJECT TITLE
EPRI / MTR
CO_2 MEMBRANE RETROFIT

BLOCK FLOW DIAGRAM

WORLEYPARSONS DWG. NO.: EPRI-MTR-0-DW-021-211-0001

REV: A



NOTES:

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ENGINEER/TECH SPECIALIST
E. MACE

DATE
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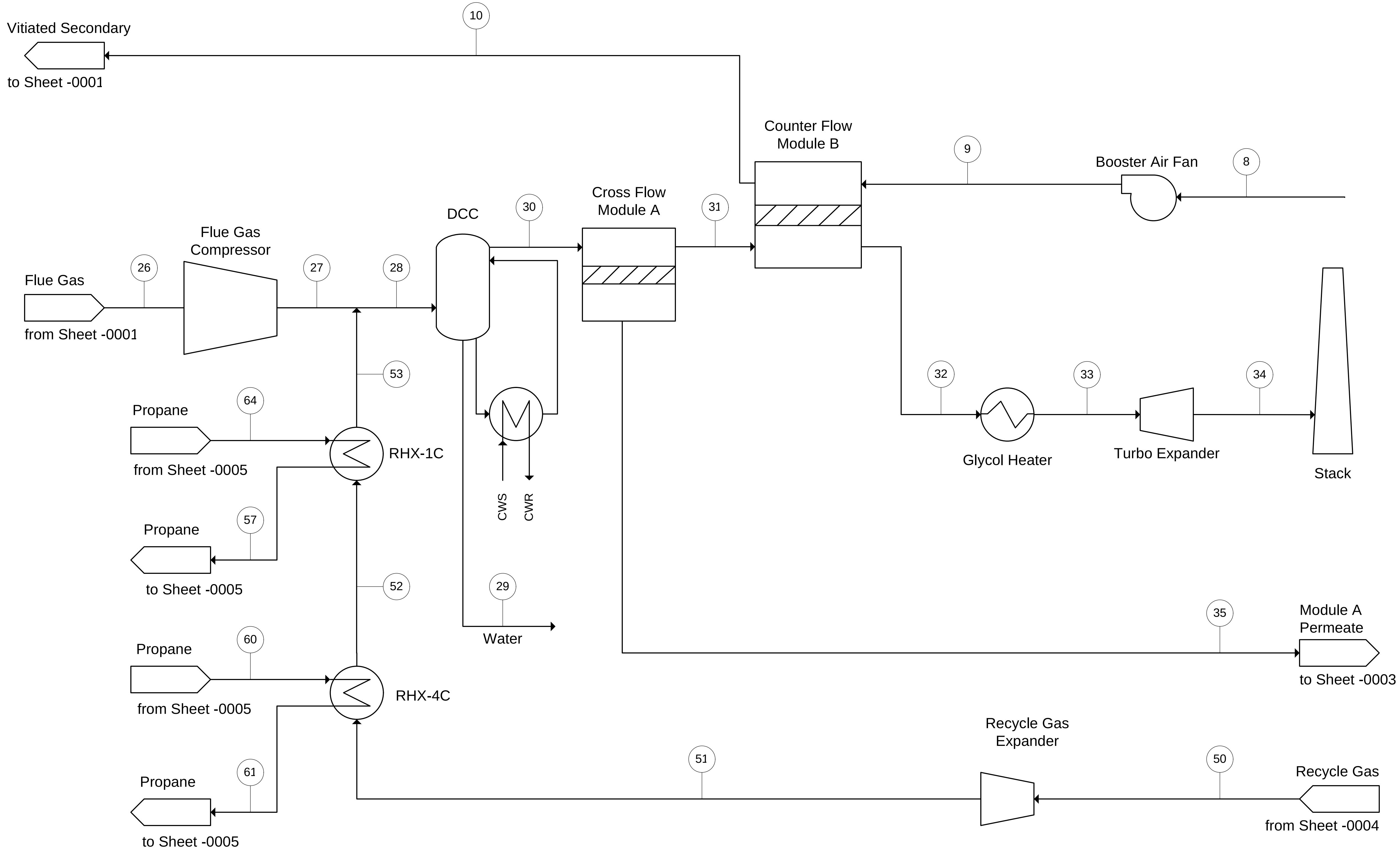
CLIENT/PROJECT TITLE
EPRI / MTR
CO_2 MEMBRANE CAPTURE

PROCESS FLOW DIAGRAM
BOILER AIR & GAS SIDE

WORLEYPARSONS DWG. NO.
EPRI-MTR-0-DW-021-305-0001

REV
A



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REV	DATE	DESCRIPTION

ENGINEER/TECH SPECIALIST E. MACE	DATE 06/10/2015
CHECKED BY	PROJECT MANAGER D. STAUFFER

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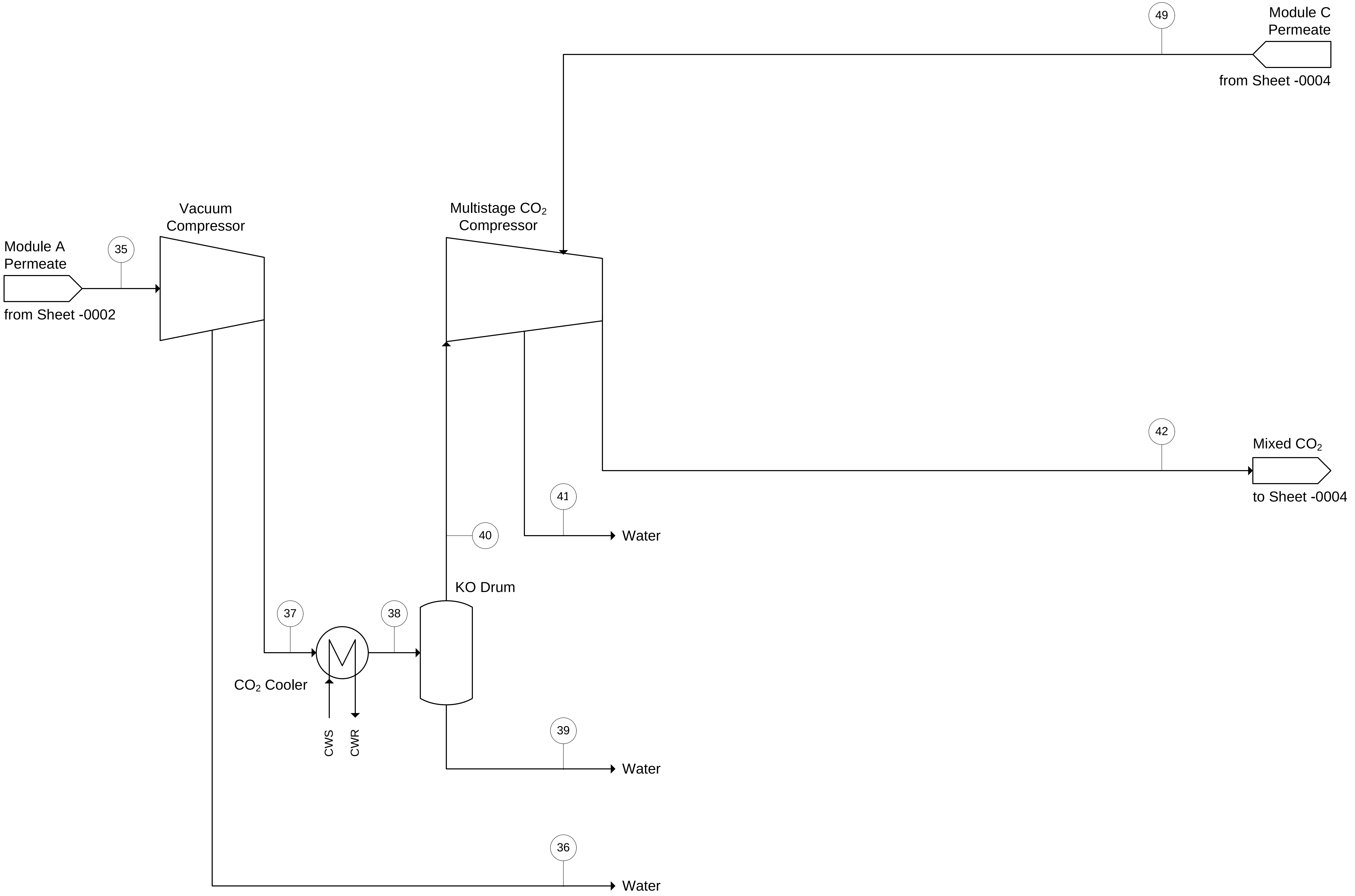
CLIENT/PROJECT TITLE
EPRI / MTR
CO_2 MEMBRANE CAPTURE

PROCESS FLOW DIAGRAM
MEMBRANE CO_2 CAPTURE

WORLEYPARSONS DWG. NO.	REV
EPRI-MTR-0-DW-021-305-0002	A



NOTES:

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ENGINEER/TECH SPECIALIST
E. MACE

DATE
06/10/2015

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PROJECT MANAGER
D. STAUFFER

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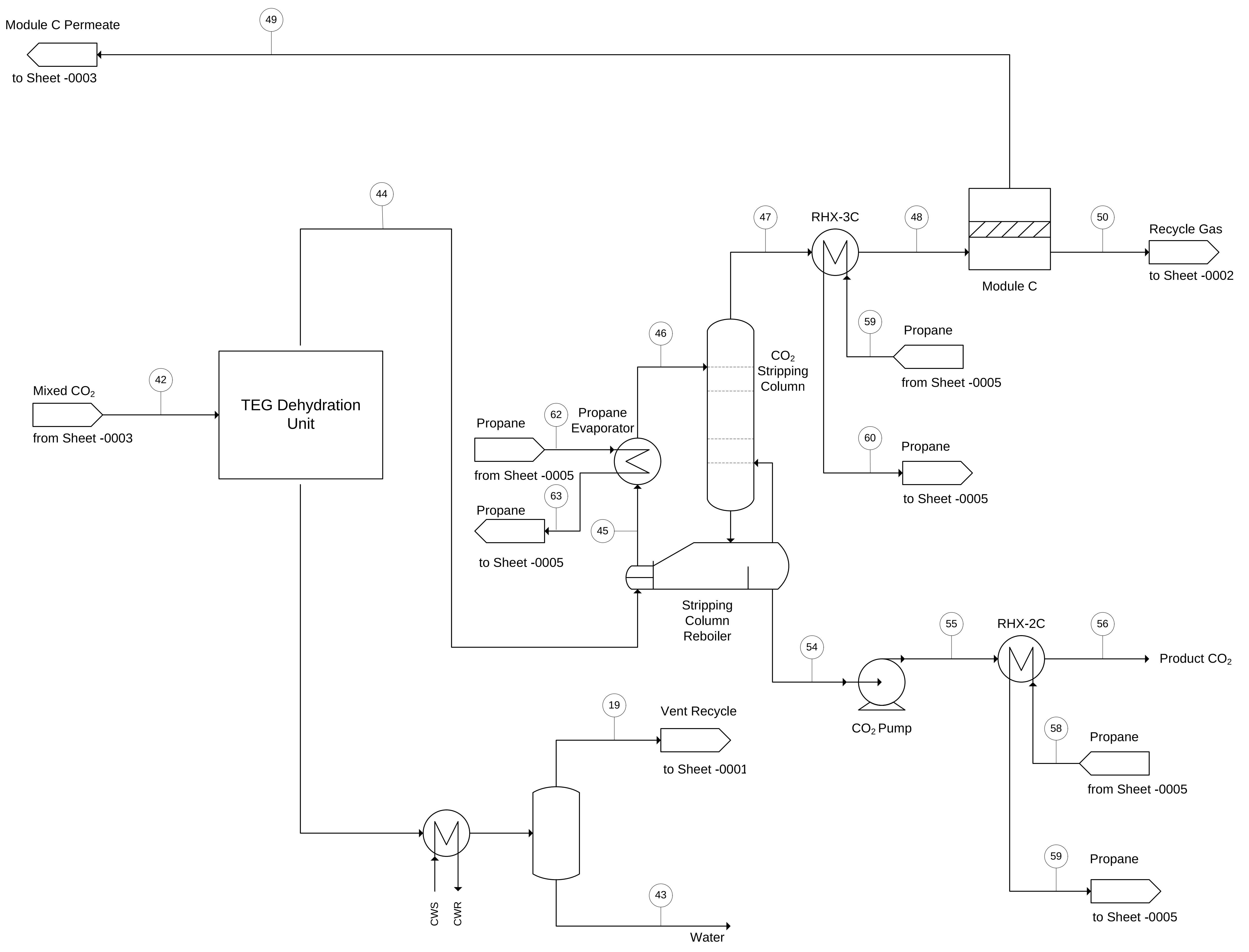
CLIENT/PROJECT TITLE
EPRI / MTR
CO_2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
CO_2 COMPRESSION

WORLEYPARSONS DWG. NO.
EPRI-MTR-0-DW-021-305-0003

REV
A



NOTES:

A	06/10/2015	INITIAL ISSUE
REV	DATE	DESCRIPTION

ENGINEER/TECH SPECIALIST E. MACE	DATE 06/10/2015
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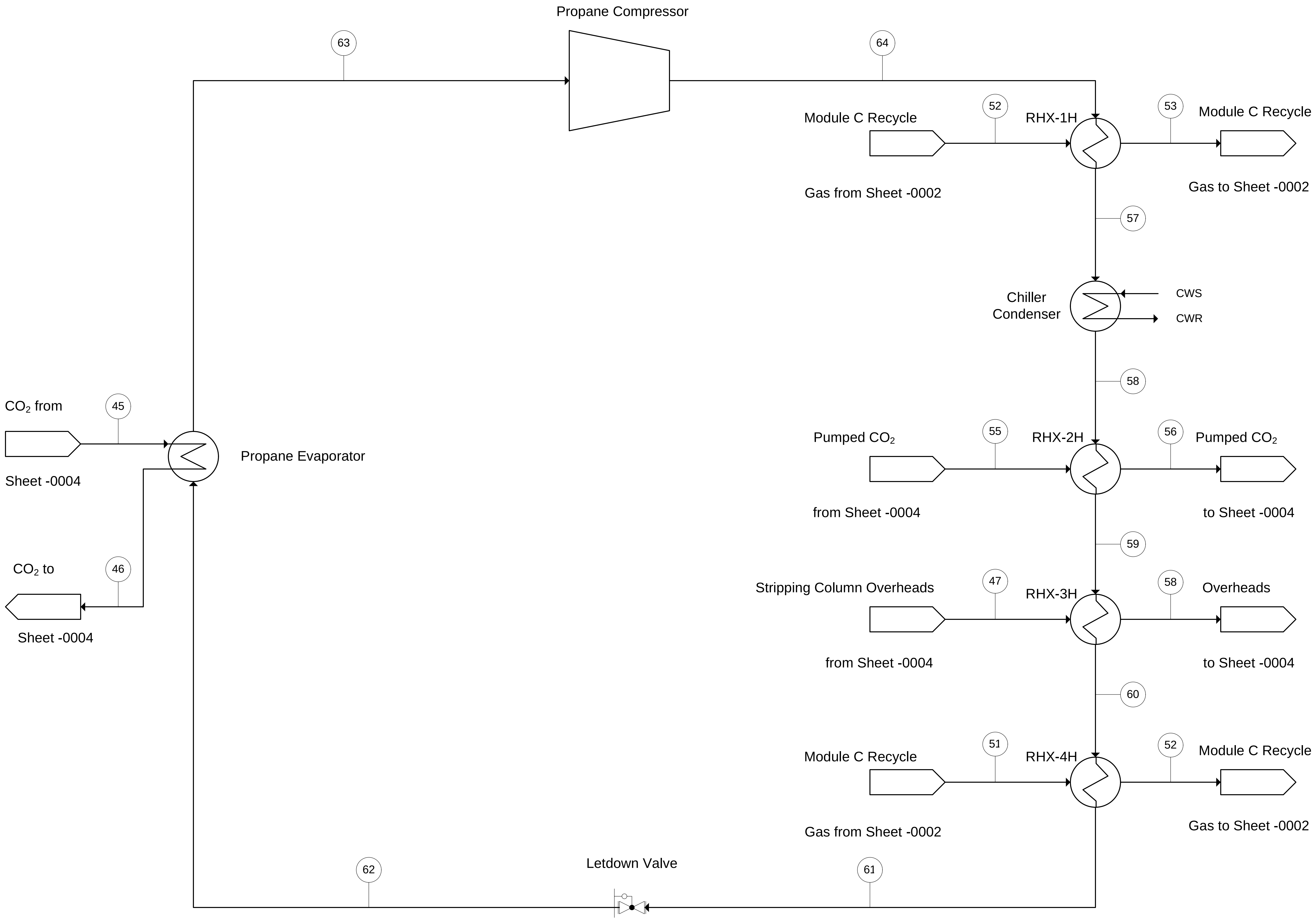
CLIENT/PROJECT TITLE
EPRI / MTR
CO_2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
CO_2 PURIFICATION

WORLEYPARSONS DWG. NO.	REV
EPRI-MTR-0-DW-021-305-0004	A



NOTES:

A	06/10/2015	INITIAL ISSUE
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ENGINEER/TECH SPECIALIST
E. MACE

DATE
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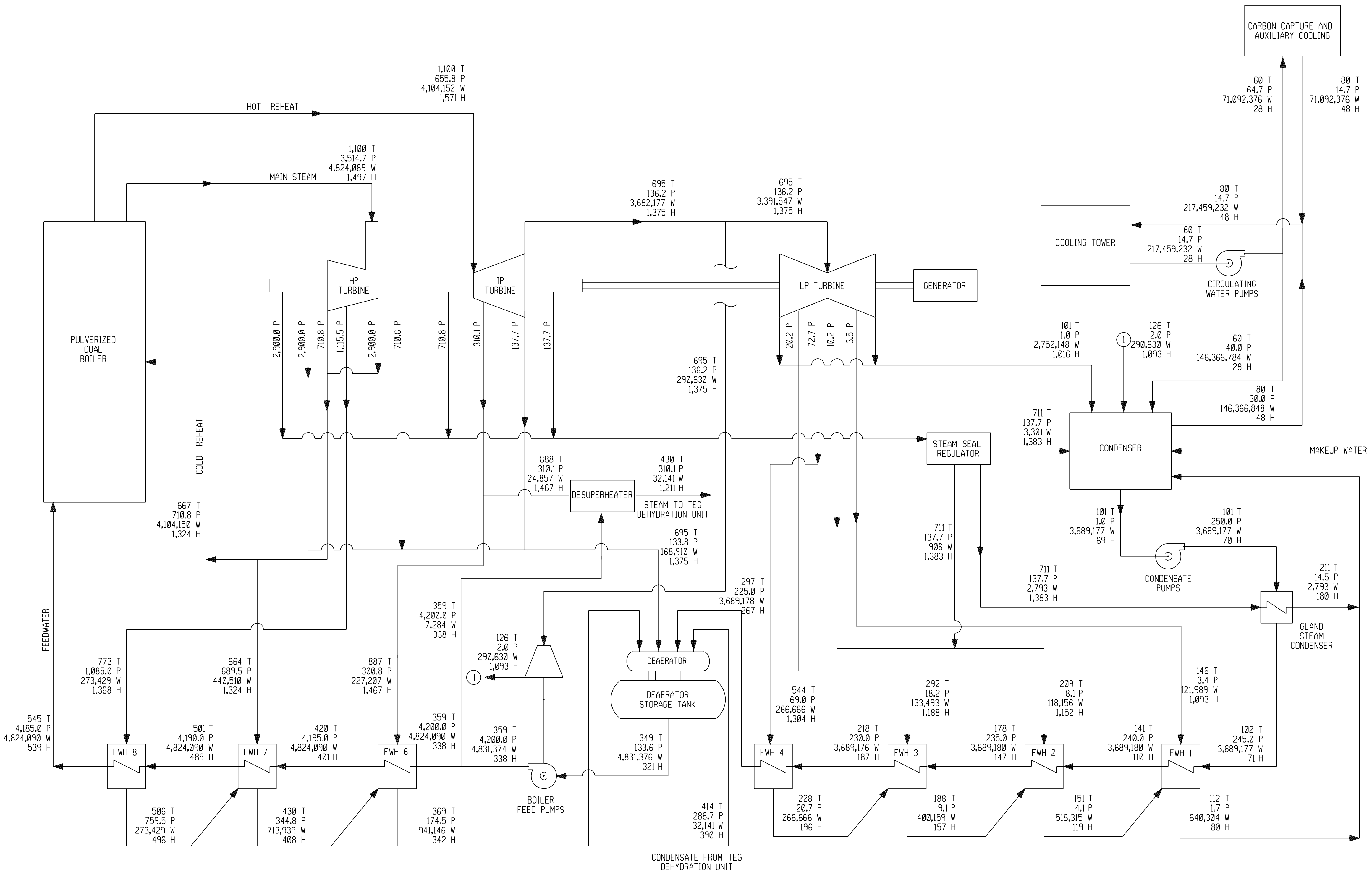
CLIENT/PROJECT TITLE
EPRI / MTR
CO_2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
REFRIGERATION UNIT

WORLEYPARSONS DWG. NO.
EPRI-MTR-0-DW-021-305-0005

REV
A

Appendix 3: Steam Cycle Heat and Mass Balance Diagram



LEGEND

───	WATER OR STEAM
T	TEMPERATURE, °F
P	PRESSURE, PSIA
W	MASS FLOW, LB/HR
H	ENTHALPY, BTU/LB
KWe	POWER, KILOWATTS ELECTRICAL

SYSTEM PERFORMANCE SUMMARY

STG OUTPUT	780,795 KW
DRY BULB	59 F
WET BULB	51.5 F
ELEVATION	0 FT
POWER FACTOR	0.9

D

PRELIMINARY STATUS	DATE	REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.
APPROVED STATUS	DATE	REPRESENTS REVIEWED AND APPROVED DESIGN. ANY PORTION MARKED "HOLD" RETAINS PRELIMINARY STATUS.

ORIGINATING PERSONNEL	PROFESSIONAL ENGINEER'S SEAL
DRAWN BY E. MACE	
CHECKED BY Y. LU	
LEAD DESIGNER	
ENGINEER/TECH SPECIALIST E. MACE	
PROJECT ENGINEERING MANAGER	
PROJECT MANAGER D. STAUFFER	
06/09/2015	ORIGINALLY PREPARED UNDER THE RESPONSIBLE SUPERVISION OF PE: __________ STATE: ______ LIC. NO.: __________ DATE: ______

WorleyParsons
resources & energy

CLIENT/PROJECT TITLE

EPRI / MTR CARBON CAPTURE PROCESS
HEAT & MASS BALANCE DIAGRAM

SUPERCRITICAL POST-COMBUSTION
MEMBRANE CAPTURE

SCALE NONE
DRAWING SIZE

WORLEYPARSONS DWG. NO. EPRI-MTR-0-HT-021-0001
REV A

ISSUED FOR REVIEW

	PROJECT MANAGER
	PROJ ENGR MANAGER
	LEAD DISC ENGINEER
	ENGINEER/ TECH. SPEC.
	LEAD DESIGNER
	CHECKED
	DRAWN
	DATE
A	REV

FAC PIPING I & C ELECT MECH CIVIL STRUC FAC PIPING I & C ELECT MECH CIVIL STRUC FAC PIPING I & C ELECT MECH CIVIL STRUC FAC PIPING I & C ELECT MECH CIVIL STRUC FAC PIPING I & C ELECT MECH CIVIL STRUC

Reuse Assessment of Water Recovered from MTR's Post-combustion, Membrane-based Process

July 2015

APPENDIX B

ABSTRACT

In addition to separating CO_2 from flue gas, MTR's post combustion, membrane-based process also separates water. The process has the potential to recover over 96% of the water from the flue gas in a commercial application. Water samples from the current 0.05 and 1.0 MWe bench and pilot scale test systems at the National Carbon Capture Center were collected and analyzed to assess the quality and potential reusability of water streams generated by the MTR membrane capture system. Based on the process configuration and water analysis results from the National Carbon Capture Center, two potential applications for water reuse were assessed—power plant cooling and makeup to boiler water and steam cycle. Since trace amounts of regulated metals such as selenium, mercury and arsenic were present in the samples and would pose a concern to plant reuse in the plant cooling system, it was determined that the recovered water would be best suited for boiler and steam cycle makeup as long as the water is treated and meets required makeup water guidelines.

CONTENTS

LIST OF FIGURES

LIST OF TABLES

1
INTRODUCTION

Background

MTR's post-combustion, membrane-based process to capture CO_2 from power plant flue gas has the added benefit of capturing water as well. Figure 1-1 is a simplified block flow diagram (BFD) of the process.

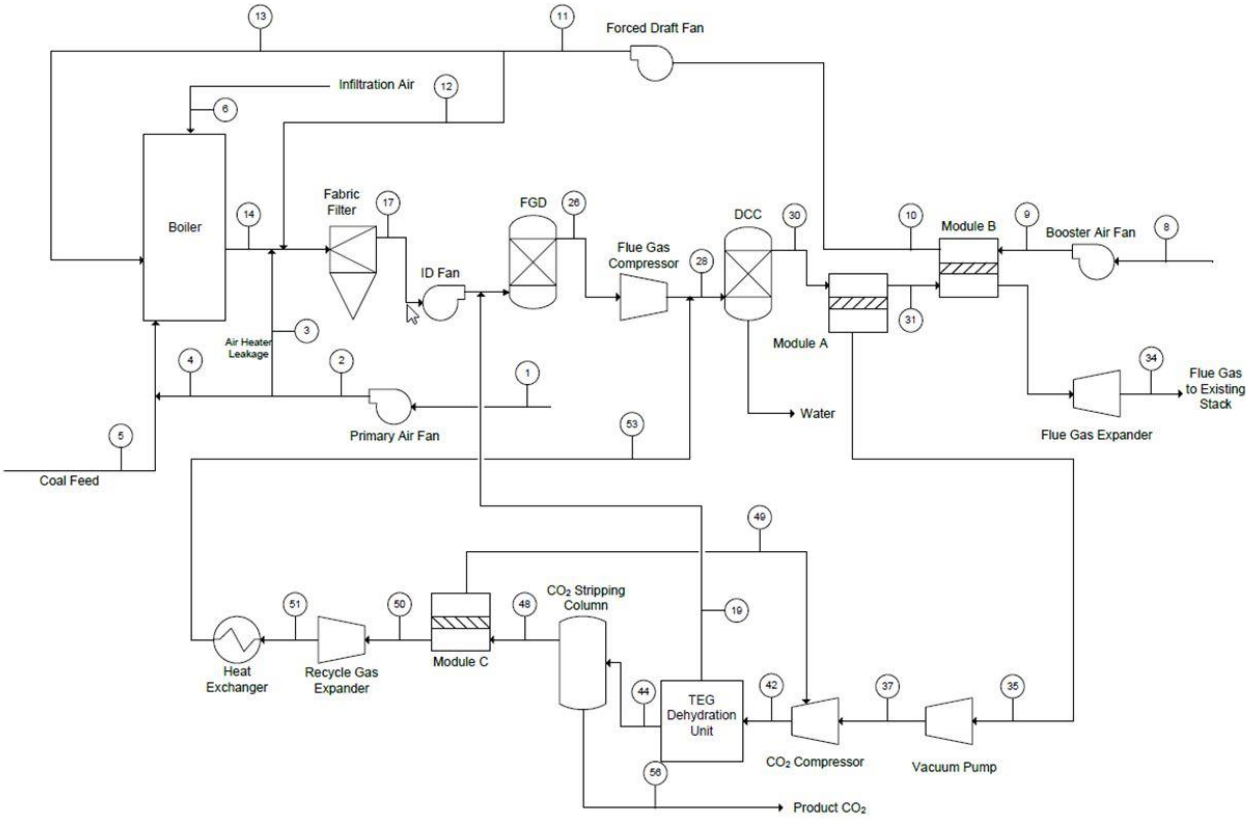


Figure 1-1
Post-Combustion CO_2 Membrane Process Block Flow Diagram

In a commercial process application, it is estimated that over 96% of the water from the flue gas exiting the flue gas desulfurization (FGD) unit can be recovered. In a full-scale implementation, a direct contact cooler placed after the FGD will remove ~85% of the water, and the cross-flow membrane (Module A on the BFD) will separate ~11% of the water from the flue gas into the permeate stream (Stream #35 on the BFD). Of the remaining moisture in the flue gas entering the air-swept membrane (Module B on the BFD), some will permeate and be recycled back to the combustion air stream to the boiler, and the balance will exit the stack. Thus, the two main areas where water can be extracted from the process are the permeate stream and the gas/water separation vessel upstream of the cross-flow membrane.

Task Objective

The objective of this task is to assess the quality and potential reusability of water streams generated by the MTR membrane capture system at full-scale through the collection and measurement of relevant samples from the current 0.05 and 1.0 MWe bench and pilot scale test systems at the National Carbon Capture Center.

Approach

Southern Research Institute has been tasked with taking water samples at various points in the CO_2 capture process. Sample bottles were filled at the sampling locations, and the samples were then sent out for analysis.

The water samples were analyzed for metals, mercury, anions and ammonia by Element One, Inc., based in Wilmington, North Carolina. Conductivity and pH measurements were conducted in house at the PC4 and PSDF Laboratories at the National Carbon Capture Center.

- For metals analysis, the samples were digested and prepared in trace metals grade nitric acid and hydrochloric acid according to SW-846 Method 3010A protocol. The prepared samples were analyzed according to EPA Method 200.8 using a PerkinElmer ELAN 6100 ICP-MS. The analytical reporting limits were 50.0 µg/L for calcium, 20.0 µg/L for sodium, magnesium, potassium and iron, and 5.0 µg/L for aluminum, arsenic, barium, and selenium.
- For mercury analysis, the samples were digested and prepared according to EPA Method 245.1 protocol. The samples were analyzed on a PerkinElmer FIMS-100 CVAA mercury analyzer. The reporting limit for mercury was 0.0002 µg/mL.
- For anion analysis, the samples were prepared and analyzed according to EPA Method 300.0 protocol. The samples were analyzed on a Metrohm 861/788 Ion Chromatograph System. The reporting limits were 0.02 µg/mL for nitrate and 0.1 µg/mL for both chloride and sulfate.
- For ammonia analysis, the samples were analyzed according to Standard Methods 4500-NH3 D (Phenate Method) protocol by UV/VIS spectroscopy. The reporting limit was 0.1 µg/mL.
- For pH analysis, the samples were analyzed according to Standard Methods 4500-H+ B protocol on an Orion Star A214 pH/ISE meter. The pH results are accurate within +/- 0.1 SU.
- For conductivity analysis, the samples were analyzed according to Standard Methods 2510 B protocol on a Myron L Company Ultrameter II. The precision of the measurement is within a range of 0.1 – 1%.

All QA/QC data including duplicate analysis RPD, spike sample recovery and second source calibration verification data were within the criteria of each respective analytical method.

It is important to note that the nitrate analyses and conductivity and pH measurements were conducted outside of the generally accepted and recommended analytical holding times because the samples were initially presented for analysis outside of those time restraints.

The results of the water analyses were then reviewed by EPRI to assess the quality of the water for potential reuse in the power plant.

2

0.05 MW BENCH SCALE UNIT

The 0.05 MWe, 1 tonne CO_2/day (1 TPD), unit entered service early in 2012 and has provided over 7,000 test hours of data in support of the design for the 1 MWe pilot unit.

Water Sampling

Water samples were taken at the following four locations to identify the species present and determine what treatment may be required before the water can be reused:

1. Compressor inlet knockout (Sample # BB00890)
   The inlet flue gas from the caustic scrubber is cooled, and this is the condensate.
2. Compressor discharge (Sample # BB00892)
   This is a sample of the cooling water discharge from the liquid ring compressor.
3. Gas/water separator (Sample # BB00893)
   The exit gas from the liquid ring compressor is cooled, and this is the condensate.
4. Vacuum pump discharge (Sample # BB00891)
   This is a sample of the cooling water discharge from the liquid ring vacuum pump.

Figure 2-1 and Figure 2-2 show a piping/instrumentation and a simplified process flow diagram, respectively that identify the water sampling locations in the process.

Figure 2-3 through Figure 2-6 are photos of the area around each sampling location.

Table 2-1 summarizes the results of the water analysis.

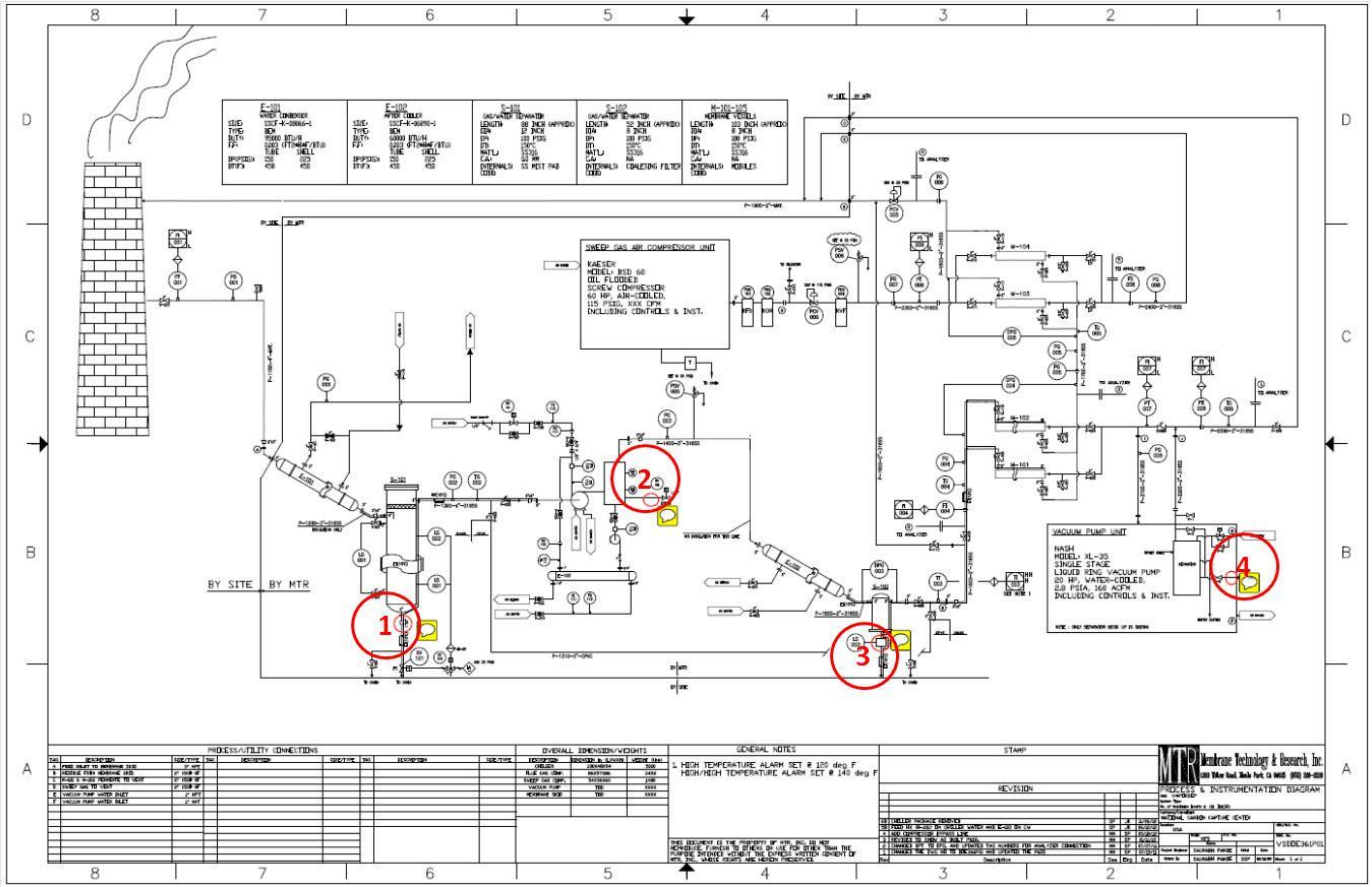


Figure 2-1
Piping & Instrumentation Diagram of 0.05 MWe Unit Showing Water Sampling Locations

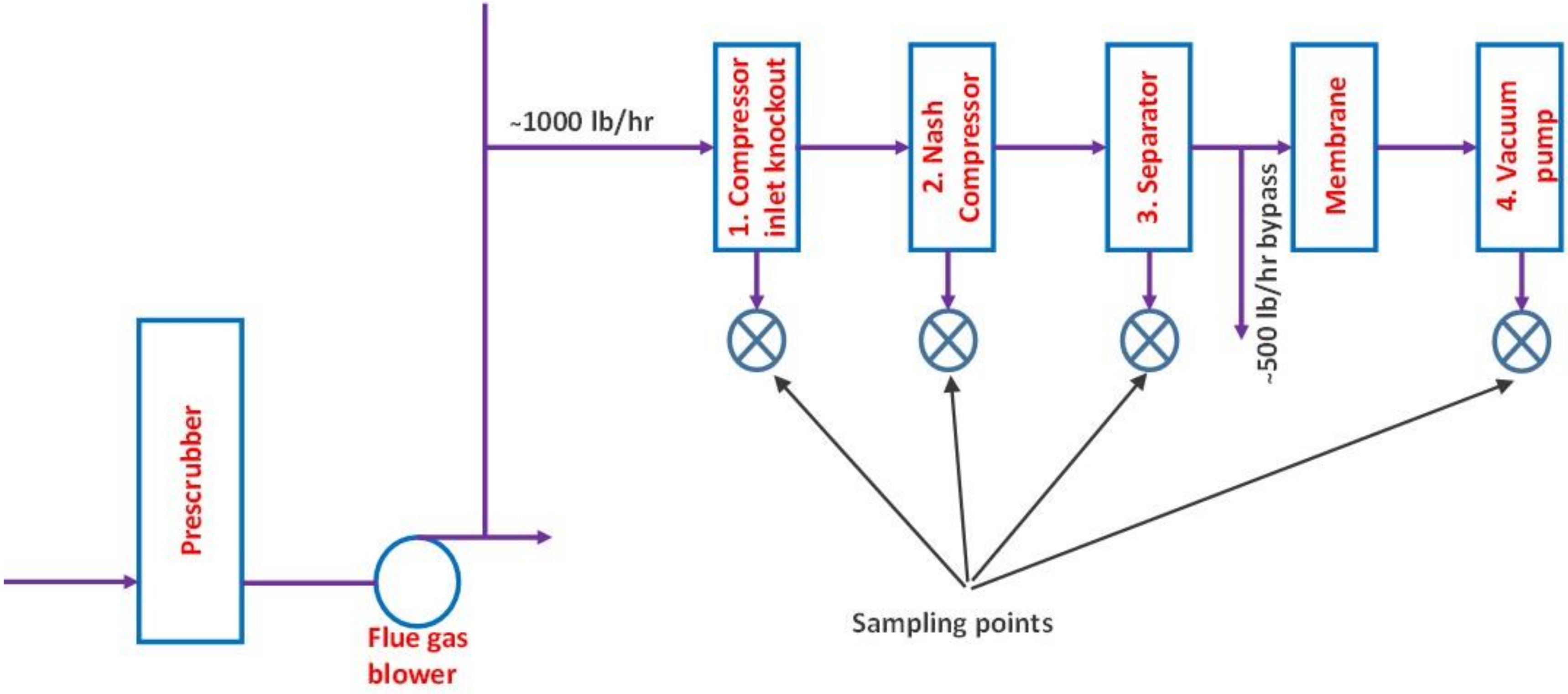


Figure 2-2
Simplified Process Flow Diagram of 0.05 MWe Unit Showing Water Sampling Locations

Figure 2-3
Photo of Compressor Inlet Knockout Sampling Location

Figure 2-4
Photo of Compressor Discharge Sampling Location

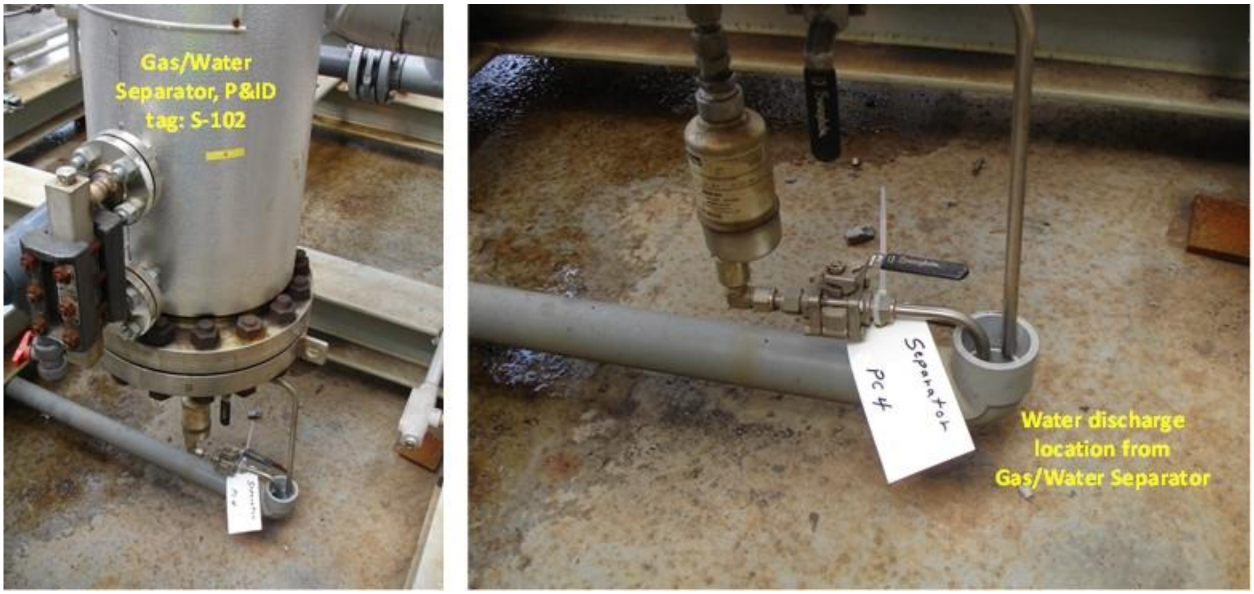


Figure 2-5
Photo of Gas/Water Separator Sampling Location

Figure 2-6
Photo of Vacuum Pump Discharge Sampling Location

Table 2-1
Water Analysis Results for 0.05 MWe Unit

Results in units of ug/mL:

SAMPLE #	Description	Collection	METALS											ANIONS		
		Date/Time	Aluminum, ug/mL	Arsenic, ug/mL	Barium, ug/mL	Calcium, ug/mL	Iron, ug/mL	Magnesium, ug/mL	Mercury, ug/mL	Potassium, ug/mL	Selenium, ug/mL	Sodium, ug/mL	Ammonia as N, ug/mL	Chloride, ug/mL	Nitrate as N, ug/mL	Sulfate, ug/mL
BB00890	COMPRESSOR INLET KNOCKOUT CONDENSATE	3/24/15 0:00	0.047	0.008	<0.005	<0.05	0.256	<0.02	0.0006	<0.02	0.068	0.048	0.259	0.201	<0.02	36.7
BB00891	VACUUM PUMP DISCHARGE CONDENSATE	3/24/15 0:00	0.166	<0.005	0.023	15.0	9.16	4.17	0.0002	1.10	<0.005	4.52	0.217	6.35	3.81	170
BB00892	COMPRESSOR DISCHARGE CONDENSATE	3/24/15 0:00	0.024	0.007	0.042	31.0	0.210	8.16	<0.0002	2.28	0.064	9.49	0.151	12.3	0.828	82.7
BB00893	SEPARATOR CONDENSATE	3/24/15 0:00	0.051	0.092	<0.005	4.96	0.626	1.48	0.0021	0.441	0.891	1.76	0.473	0.359	0.334	473

Results in units of lb/hr:

SAMPLE #	Description	Collection	METALS											ANIONS		
		Date/Time	Aluminum, lb/hr	Arsenic, lb/hr	Barium, lb/hr	Calcium, lb/hr	Iron, lb/hr	Magnesium, lb/hr	Mercury, lb/hr	Potassium, lb/hr	Selenium, lb/hr	Sodium, lb/hr	Ammonia as N, lb/hr	Chloride, lb/hr	Nitrate as N, lb/hr	Sulfate, lb/hr
BB00890	COMPRESSOR INLET KNOCKOUT CONDENSATE	3/24/15 0:00	9.3839E-06	1.6E-06	< 9.98E-07	< 9.98E-06	5.11E-05	< 3.99E-06	1.2E-07	< 3.99E-06	1.358E-05	9.58E-06	5.1711E-05	4.01E-05	< 3.99E-06	0.007327
BB00891	VACUUM PUMP DISCHARGE CONDENSATE	3/24/15 0:00	2.48991E-05	< 7.50E-07	3.45E-06	0.0	0.001374	0.00062548	3E-08	0.00	< 7.50E-07	0.000678	3.2549E-05	0.000952	0.000571	0.025499
BB00892	COMPRESSOR DISCHARGE CONDENSATE	3/24/15 0:00	2.8921E-06	8.44E-07	5.06E-06	0.0	0.000	0.00098332	< 2.41E-08	0.0002747	7.712E-06	0.001144	1.8196E-05	0.001482	9.98E-05	0.009966
BB00893	SEPARATOR CONDENSATE	3/24/15 0:00	6.87038E-06	1.24E-05	< 6.74E-07	0.000668	8.43E-05	0.00019938	2.83E-07	5.941E-05	0.00012	0.000237	6.3719E-05	4.84E-05	4.5E-05	0.063719

Water recovered during sampling with units of mL:

BB00890	COMPRESSOR INLET KNOCKOUT CONDENSATE	3/24/15 0:00	~500 mL
BB00891	VACUUM PUMP DISCHARGE CONDENSATE	3/24/15 0:00	~500 mL
BB00892	COMPRESSOR DISCHARGE CONDENSATE	3/24/15 0:00	~500 mL
BB00893	SEPARATOR CONDENSATE	3/24/15 0:00	~500 mL

3

1 MW PILOT SCALE UNIT

Testing of the 1 MWe pilot unit included an assessment of the water recovery potential from the knockout/separator vessels and the product CO_2 stream.

Water Sampling

Water samples were taken at the following locations to identify the species present and determine what treatment may be required before the water can be reused:

1. Cooler condenser following the compressor (Sample # BB00959)
2. Liquid from knockout vessel (Sample # BB00960)
3. Permeate after the cross-flow membrane
4. Retentate after the cross-flow membrane

Figure 3-1 and Figure 3-2 show a piping/instrumentation and a simplified process flow diagram, respectively that identify the water sampling locations in the process.

Figure 3-3 through Figure 3-5 are photos of the area around each sampling location.

Note that at the retentate sampling location, nothing condensed out indicating minimal or no water present during sampling.

For the permeate sampling location, water sampling results were not available at the time this report was written.

Table 3-1 summarizes the results of the water analysis.

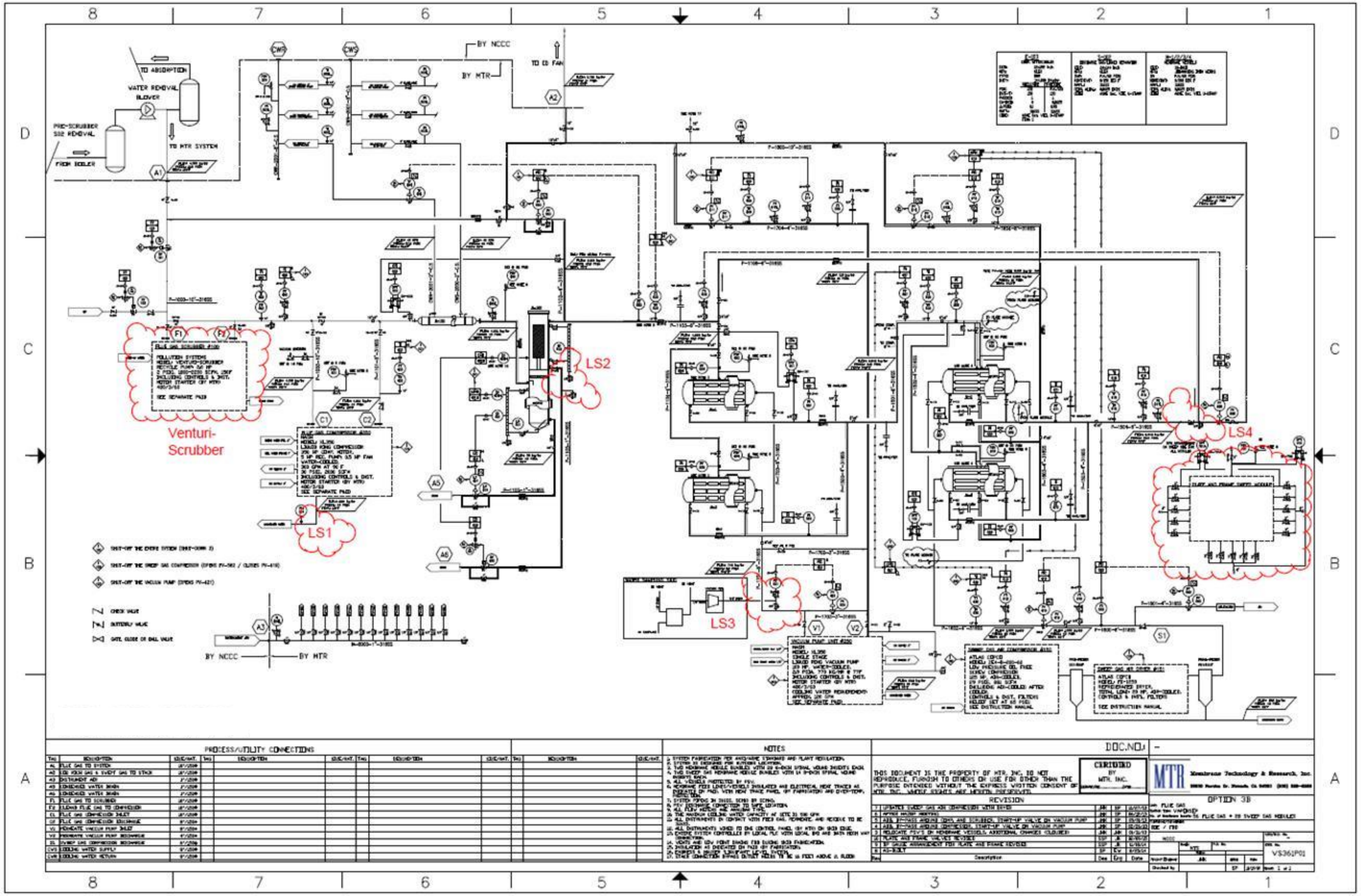


Figure 3-1
Piping & Instrumentation Diagram of 1 MWe Unit Showing Water Sampling Locations

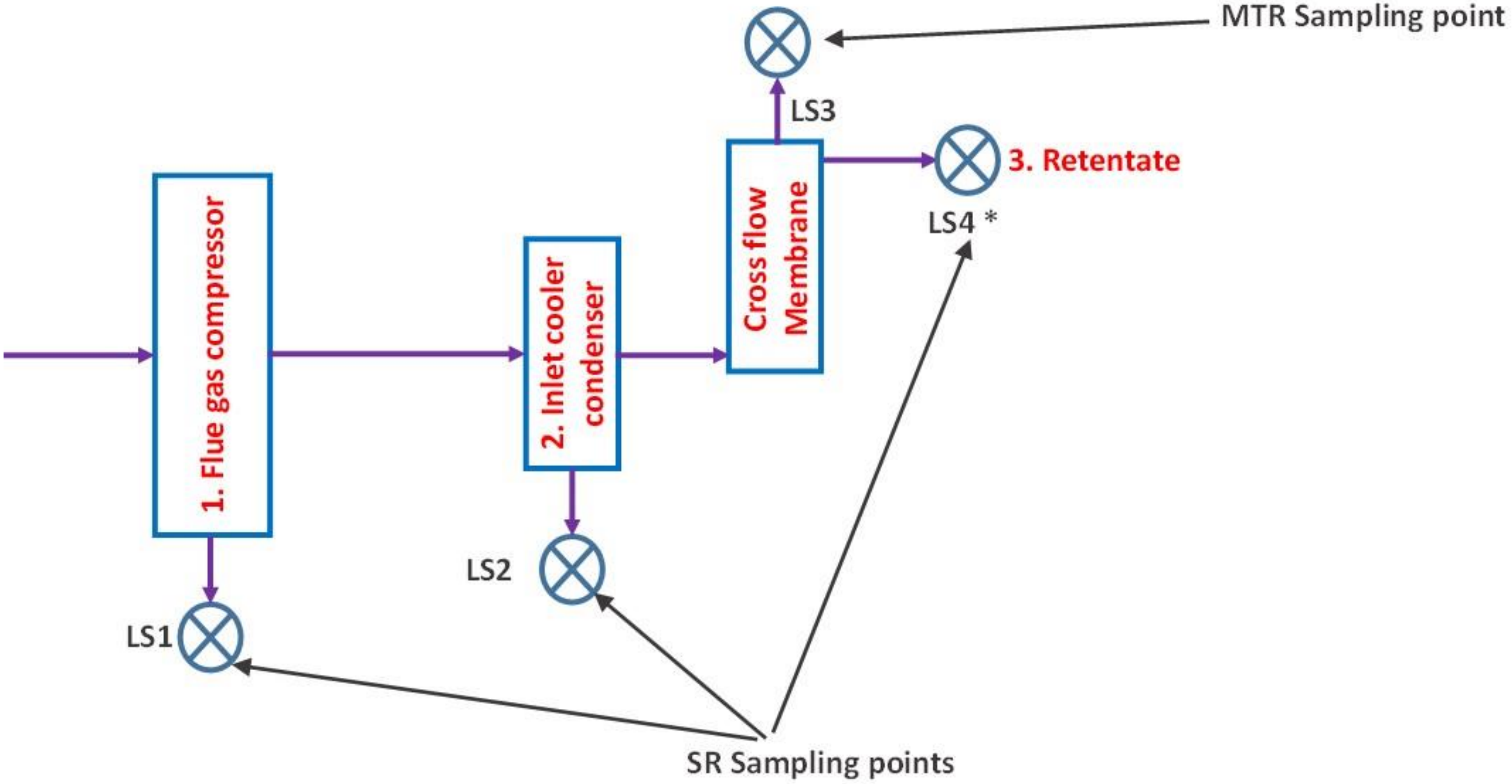


Figure 3-2
Simplified Process Flow Diagram of 1 MWe Unit Showing Water Sampling Locations

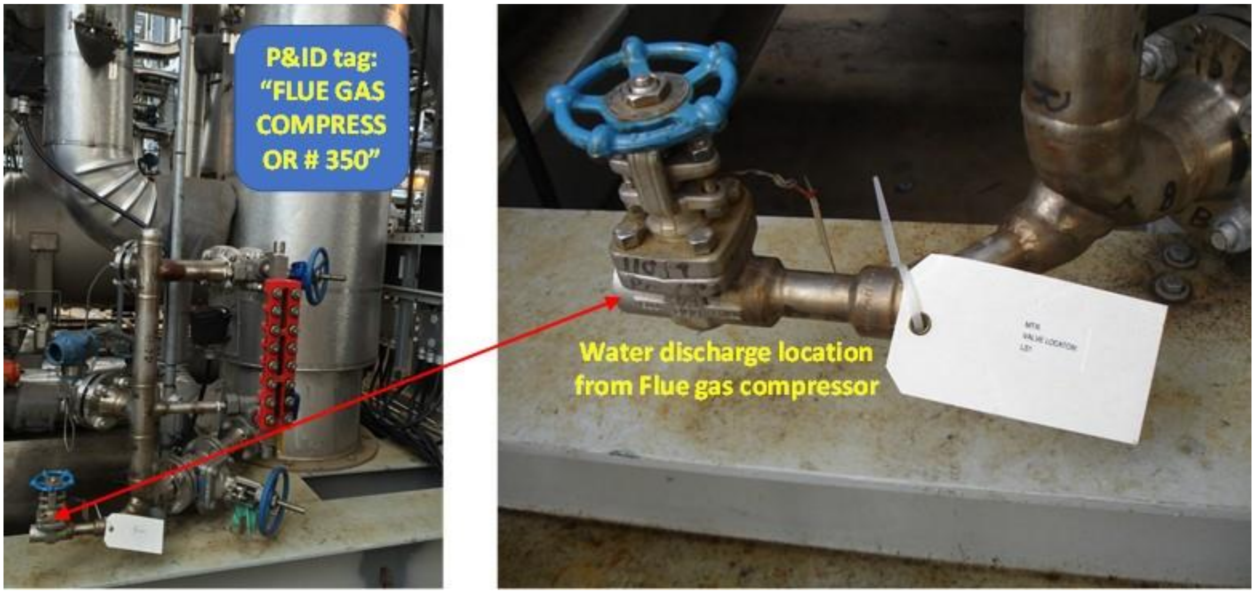


Figure 3-3
Photo of Cooler Condenser following the Compressor Sampling Location

Figure 3-4
Photo of Knockout before Entering Membrane Sampling Location

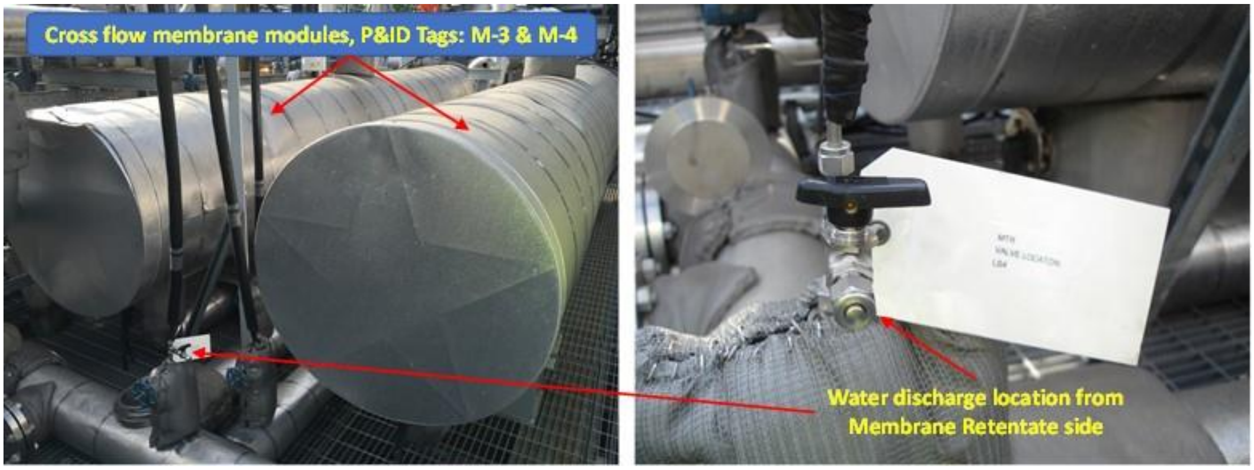


Figure 3-5
Photo of Gas/Water Separator Sampling Location

Table 3-1
Water Analysis Results for 1 MWe Unit

Results in units of ug/mL:

SAMPLE #	Description	Collection	METALS											ANIONS		
		Date/Time	Aluminum, ug/mL	Arsenic, ug/mL	Barium, ug/mL	Calcium, ug/mL	Iron, ug/mL	Magnesium, ug/mL	Mercury, ug/mL	Potassium, ug/mL	Selenium, ug/mL	Sodium, ug/mL	Ammonia as N, ug/mL	Chloride, ug/mL	Nitrate as N, ug/mL	Sulfate, ug/mL
BB00959	CONDENSATE AT LS1	5/14/15 11:40	0.020	<0.005	<0.005	0.261	0.157	0.035	<0.0002	0.030	0.011	0.411	<0.168	<0.1	0.358	8.62
BB00960	CONDENSATE AT LS2	5/14/15 11:40	0.068	0.050	0.019	10.5	0.881	2.626	<0.0002	0.744	0.609	3.886	0.406	4.42	0.151	208

Results in units of lb/hr:

SAMPLE #	Description	Collection	METALS											ANIONS		
		Date/Time	Aluminum, lb/hr	Arsenic, lb/hr	Barium, lb/hr	Calcium, lb/hr	Iron, lb/hr	Magnesium, lb/hr	Mercury, lb/hr	Potassium, lb/hr	Selenium, lb/hr	Sodium, lb/hr	Ammonia as N, lb/hr	Chloride, lb/hr	Nitrate as N, lb/hr	Sulfate, lb/hr
BB00959	CONDENSATE AT LS1	5/14/15 11:40	0.000	< 1.38E-05	< 1.38E-05	0.000722	0.000434	9.6816E-05	< 5.53E-07	0.000	3.043E-05	0.001137	< 4.65E-04	< 2.77E-04	0.00099	0.023844
BB00960	CONDENSATE AT LS2	5/14/15 11:40	7.52698E-05	0.000	2.1E-05	0.011623	0.000975	0.00290674	< 2.21E-07	0.0008235	0.0006741	0.004301	0.0004494	0.004893	0.000167	0.230237

Physical Characteristics:

SAMPLE #	Description	Collection	pH, SU	Conductivity, uS/cm
BB00959	CONDENSATE AT LS1	5/14/2015 11:40	3.67	93.0
BB00960	CONDENSATE AT LS2	5/14/2015 11:40	2.49	1520

Water recovered during sampling with units of mL:

BB00959	CONDENSATE AT LS1	5/14/2015 11:40	550 mL
BB00960	CONDENSATE AT LS2	5/14/2015 11:40	550 mL

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WATER REUSE ASSESSMENT

MTR's post-combustion membrane process to capture CO_2 from power plant flue gas has the added benefit of capturing water as well. However, the recovered water via condensate through the process also captures contaminates in the flue gas which will affect water reuse applications in the power plant.

Two potential applications for water reuse are assessed and described in the following sections.

Makeup to Boiler Water and Steam Cycle

If the recovered water would be used for makeup to the boiler water and steam cycle, the purity of water and steam is central to ensuring fossil plant component availability and reliability. EPRI has published comprehensive guidelines for drum and once-through units, and these guidelines provide information on the application of several chemical treatment strategies such as all-volatile treatment (AVT), oxygenated treatment (OT), phosphate treatment (PT), and caustic treatment (CT). The guidelines help operators reduce corrosion and deposition and thereby achieve significant operation and maintenance cost reductions and greater unit availability. These guidelines have been developed to address the serious corrosion and deposition problems that have been experienced in fossil power plants. These problems include chemistry-influenced boiler tube failures, turbine corrosion, and deposition and flow-accelerated corrosion.

The recovered water from the MTR process is not suitable for boiler water makeup, and several constituents exceed the recommended purity standards in the published EPRI guidelines. In particular, anions such as chloride and sulfate are well above purity standards of less than two parts per billion (ppb) and removal of these constituents by ion exchange demineralizer resin applications would be essential to protect the boiler and steam turbine from corrosion. Though this polishing step is often routine and well understood in the power plant makeup water treatment system, there could be a concern from several non-typical potential boiler water contaminates in this recovered water such as selenium, arsenic, and mercury. The analysis in this study does not include speciation of these metals, and thus it is difficult to assess their removal efficiency in the ion-exchange process.

These metals may also be colloidal in nature and therefore, it would also be highly probable that if the recovered water was to be used for boiler and steam cycle makeup, the water would also need to be processed by advanced filtration applications such as ultra-filtration (UF) and/or nano-filtration (NF) and then followed by reverse osmosis (RO) or a similar membrane treatment prior to processing by demineralizer ion-exchange.

It is important to note that each of these processes only will separate contaminates from the processed water, and the reject stream from both filtration (backwash) and brine (RO reject) will have to be treated. This reject or wastewater stream would require treatment as it has the potential to exceed industry regulations, especially concerning the metals.

If the metals were not removed, the fate of selenium, mercury, and arsenic are not understood in the boiler and steam cycle as these are not typical contaminates in the makeup water systems currently in practice. There would be a concern of volatilization of these metals at the extremely high temperatures and pressures experienced in the boiler and steam cycle and the negative effect they would pose to the turbine and boiler system, including the chemistry at the deaerator vents.

Power Plant Cooling

Another approach for the recovered water could be for cooling. However, this also may present several concerns, since many cooling water systems at the power plant, specifically cooling towers, are concentrating systems. As water evaporates as part of the cooling process, dissolved species will remain in the bulk water. This evaporation mechanism of pure water in the cooling tower plume therefore results in a concentrated cooling water, and the cooling water in the recirculating system will be several times concentrated than the original make up water. This is known as cycles of concentration (COC), and the calculation of the COC is a routine practice in power plant operations. Cycles of concentration varies throughout the industry and is dependent upon many factors including makeup water chemistry, water availability, heat load and discharge (blowdown) regulatory limits. The range of COC can be from 2x or 3x to sometimes as high as 15x to 20x. As such, the presence of regulated metals such as selenium, mercury and arsenic would pose a concern to plant reuse in the cooling system. It is also observed that the recovered water from the MTR process would be a very small contribution to the overall requirements for plant cooling needs and therefore an analysis of minimal water recovery benefits to potential environmental risk would be recommended.

Reuse Considerations

Water reclamation and reuse applications are frequently employed in power plants as part of the water balance program. Final permeate water quality of the membrane process to capture CO_2 from power plant flue still needs to be determined, as the samples in this study show some contaminates in the flue gas which will affect water reuse applications in the power plant. Given the potential of flue gas borne contaminates, even in trace amounts and the overall volume of the recovered water having a small impact on the thousand gallons per minute required for cooling; it is most likely that the recovered water would be best suited for boiler and steam cycle makeup as long as the water meets required make up water guidelines. To accomplish that effort, it is highly probable that the recovered water from the MTR membrane process would require treatment from membrane based technology, such as micro or ultra-filtration and/or reverse osmosis, followed by ion exchange via demineralization.