TITLE PAGE

Report Title:	**Electrochemical Membrane for Carbon Dioxide Capture and Power Generation**
Type of Report:	Final Technical Report
Reporting Period Start Date:	October 1, 2011
Reporting Period End Date:	May 31, 2016
Principal Author:	Hossein Ghezel-Ayagh
Date Report Issued	December 21, 2017, Revision 1
DOE Award No.:	DE-FE0007634
Name & Address of Submitter:	FuelCell Energy, Inc. 3 Great Pasture Road Danbury, CT 06813
Subcontractors:	AECOM Corporation 9400 Amberglen Blvd. Austin, TX 78729 Pacific Northwest National Laboratory 902 Battelle Blvd., Richland, WA 99352

DISCLAIMER

ABSTRACT

FuelCell Energy, Inc. (FCE), in collaboration with AECOM Corporation (formerly URS Corporation) and Pacific Northwest National Laboratory, has been developing a novel Combined Electric Power and Carbon-dioxide Separation (CEPACS) system. The CEPACS system is based on electrochemical membrane (ECM) technology derived from FCE's carbonate fuel cell products featuring internal (methane steam) reforming and carrying the trade name of Direct FuelCell®. The unique chemistry of carbonate fuel cells offers an innovative approach for separation of CO_2 from existing fossil-fuel power plant exhaust streams (flue gases). The ECM-based CEPACS system has the potential to become a transformational CO_2-separation technology by working as two devices in one: it separates the CO_2 from the exhaust of other plants such as an existing coal-fired plant and simultaneously produces clean electric power at high efficiency using a supplementary fuel.

The development effort was carried out under the U.S. Department of Energy (DOE) cooperative agreement DE-FE0007634. The overall objective of this project was to successfully demonstrate the ability of FCE's ECM-based CEPACS system technology to separate ≥90% of the CO_2 from a simulated Pulverized Coal (PC) power plant flue gas stream and to compress the captured CO_2 to a state that can be easily transported for sequestration or beneficial use. In addition, a key objective was to show, through the technical and economic feasibility study and bench scale testing, that the ECM-based CEPACS system is an economical alternative for CO_2 capture in PC power plants, and that it meets DOE's objective related to the incremental cost of electricity (COE) for post-combustion CO_2 capture (no more than 35% increase in COE).

The project was performed in three budget periods (BP). The specific objective for BP1 was to complete the Preliminary Technical and Economic Feasibility Study. The T&EF study was based on the carbon capture system size suitable for a reference 550 MW PC power plant. The specific objectives for BP2 were to perform (flue gas) contaminant effect evaluation tests, small area membrane tests using clean simulated flue gas, design a flue gas pretreatment system for processing of the gas feed to ECM, update the Technical & Economic Feasibility Study (T&EFS) incorporating results of contaminant effect tests and small area membrane tests, and to prepare a test facility for bench scale testing. The specific objectives for BP3 were to perform bench scale testing (parametric and long-duration testing) of a 11.7 m^2 ECM-based CO_2 capture, purification and compression system, and update (as final) the Technical and Economic Feasibility Study. In addition, an Environmental Health and Safety evaluation (assessment) of the ECM technology was included. This final technical report presents the progress made under the project.

TABLE OF CONTENTS

LIST OF FIGURES

LIST OF TABLES

EXECUTIVE SUMMARY

FuelCell Energy, Inc. (FCE), in collaboration with AECOM Corporation (formerly URS Corporation) and Pacific Northwest National Laboratory (PNNL), has been developing a novel Combined Electric Power (generation) and Carbon-dioxide Separation (CEPACS) system. The CEPACS system is based on electrochemical membrane (ECM) technology derived from FCE's carbonate fuel cell products featuring internal (methane steam) reforming and carrying the trade name of Direct FuelCell® (DFC®). The unique chemistry of carbonate fuel cells offers an innovative approach for separation of CO_2 from existing fossil-fuel power plant exhaust streams (flue gases). The ECM-based CEPACS system separates the CO_2 from the exhaust of other plants such as an existing coal-fired plant and simultaneously produces clean electric power at high efficiency using a supplementary fuel. During the project period, significant progress was made in the areas of Technical and Economic Feasibility Study of the CEPACS system, Pulverized Coal (PC) plant flue gas contaminant Effect (on ECM) evaluation, membrane (ECM) performance characterization, Balance-of-plant component technology gap evaluation, and bench-scale testing of ECM stack and CEPACS system. A brief summary for each follows.

Preliminary Technical and Economic Feasibility Study:
A CEPACS plant utilizing ECM technology was designed for a reference 550 MW (net AC) PC Rankine Cycle Plant to capture and compress CO_2 from the PC plant flue gas. The PC plant design specified in "Case 9" of the U.S. Department of Energy - National Energy Technology Laboratory (DOE - NETL) report "*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2" wa*s used as a reference. The system configuration, simulations and analyses were performed using CHEMCAD process simulation software to guide the conceptual design of the CEPACS plant. The performance assessment included estimation of the parasitic power consumption for >90% CO_2 capture and compression, and the efficiency impact on the PC plant. System reconfiguration to improve thermal integration, and optimization studies with respect to the operating parameters to increase plant efficiency and reduce cost, were conducted. ECM parameters such as current density, fuel utilization, cell performance (voltage) level and the number of ECM modules/plant were varied for the optimization. An equipment list and ECM module layout were prepared to facilitate the economic analysis. The CEPACS plant contained a total of 500 ECM modules. Vendor quotes including the equipment design information and cost were solicited. The economic feasibility study included estimation of CEPACS plant capital cost and cost of electricity (COE) analyses. The DOE-NETL document: 'Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance', dated April 2011, DOE/NETL-2011/1455 was used as a guideline for the study/analysis. The study was specifically focused on developing technical and economic comparisons to Cases 9 and 10 of the DOE/NETL report.

While the ECM-based CEPACS system for the 550 MW PC plant captures 90% of CO_2 from the flue gas, it generates additional (net AC) power after compensating for the auxiliary power requirements of CO_2 capture and compression. The net electrical efficiency of the PC plant equipped with a CEPACS system (for CO_2 capture) was estimated to be 39.8% (based on higher heating values of coal and natural gas fuels used by PC plant and CEPACS plant, respectively). In addition to the baseline ECM case above (Case 3), two alternative cases were studied. One case (Case 4) evaluated the performance and cost of a natural gas combined cycle (NGCC) plant integrated with the 550 MW PC plant, with the combined plant featuring Econamine-based CO_2 capture. This provided a comparison with the ECM-based CEPACS case as in this study the CEPACS system used pipeline natural gas as a supplementary fuel. The other alternative case (Case 5) considered a smaller PC plant integrated with a CEPACS

plant for >90% CO_2 capture and compression. The total net AC output of the combined plant was sized to be ~550 MW (comparable to Cases 9 and 10 of the DOE/NETL report). Table ES-1 shows the cost, performance, and environmental profile summary for all cases. Cases 1 and 2 in this study correspond to Cases 9 and 10 of the DOE/NETL report.

Table ES-1. Cost and Performance Summary for System Cases Studied

	Case 1 PC	Case 2 PC w/ Amine CO_2 cap	Case 3 PC + CEPACS	Case 4 PC + NGCC w/ Amine CO_2 cap	Case 5 ½(PC + CEPACS)
PERFORMANCE					
CO_2 Capture	0%	90%	90%	90%	90%
Gross Power Output (kWe)	582,600	672,700	1,106,140	1,183,700	553,070
Auxiliary Power Requirement (kWe)	32,580	122,740	114,968	160,170	57,484
Net Power Output (kWe)	550,020	549,960	991,172	1,023,530	495,586
Coal Flow rate (lb/hr)	437,378	614,994	437,378	614,994	218,689
Natural Gas Flow rate (lb/hr)	N/A	N/A	150,756	167,333	75,378
HHV Thermal Input (kWth)	1,495,379	2,102,643	2,489,282	3,208,455	1,244,641
Net Plant HHV Efficiency (%)	36.80%	26.20%	39.82%	31.9%	39.82%
Net Plant HHV Heat Rate (Btu/kWh)	9,277	13,046	9,041	10,696	9,041
Raw Water Withdrawal (gpm/MWnet)	10.7	20.4	8.9	14.8	8.9
Process Water Discharge (gpm/MWnet)	2.2	4.7	2.4	3.5	2.4
Raw Water Consumption (gpm/MWnet)	8.5	15.7	7.0	11.4	7.0
CO_2 Emissions (lb/MMBtu)	204	20	12	18	12
CO_2 Emissions (lb/MWhgross)	1,783	217	94	161	94
CO_2 Emissions (lb/MWhnet)	1,888	266	104	186.4	104
SO_2 Emissions (lb/MMBtu)	0.0858	0.0017	0.0006	0.0012	0.0006
SO_2 Emissions (lb/MWhgross)	0.7515	0.0176	0.0044	0.0122	0.0044
NOx Emissions (lb/MMBtu)	0.07	0.07	0.0144	0.0511	0.0144
NOx Emissions (lb/MWhgross)	0.613	0.747	0.1107	0.5374	0.1107
PM Emissions (lb/MMBtu)	0.013	0.013	0.0039	0.0090	0.0039
PM Emissions (lb/MWhgross)	0.114	0.139	0.0302	0.0963	0.0302
Hg Emissions (lb/TBtu)	1.143	1.143	0.6865	0.7922	0.6865
Hg Emissions (lb/MWhgross)	1.00E-05	1.22E-05	5.27E-06	8.46E-06	5.27E-06
COST					
Total Plant Cost (2007$/kW)	1,622	2,942	1,803	1,971	2,338
Total Overnight Cost (2007$/kW)	1,996	3,610	2,201	2,450	2,848
Bare Erected Cost	1,317	2,255	1,419	1,492	1,845
Home Office Expenses	124	213	133	140	173
Project Contingency	182	369	217	255	279
Process Contingency	0	105	34	84	41
Owner's Costs	374	667	398	479	510
Total Overnight Cost (2007$ x 1,000)	1,098,124	1,985,432	2,181,683	2,508,760	1,411,584
Total As Spent Capital (2007$/kW)	2,264	4,115	2,509	2,793	3,247
CO_2 TS&M Costs	0.0	5.8	2.2	4.6	2.2
Fuel Costs	15.2	21.3	30.7	35.6	30.7
Variable Costs	5	9.2	6.4	6.1	7.7
Fixed Costs	7.8	13.1	4.6	9.6	6.6
Capital Costs	31.2	60.2	36.7	40.8	47.4
COE (mills/kWh, 2007$)	**59.4**	**109.6**	**80.5**	**96.8**	**94.5**
Incremental COE, Case 1 Basis		84.3%	35.5%	62.9%	59.1%
LCOE (mills/kWh, 2007$)	75.3	139.0	102.0	122.7	119.8

Overall, Case 3 (CEPACS-equipped PC plant) has the potential to meet DOE's objective related to the incremental cost of electricity (COE) for post-combustion CO_2 capture (no more than 35% increase in COE). Case 3 also offers ~3 percentage point increase in PC plant efficiency as compared to the competing technology (Case 2, amine-based CO_2 capture) reducing the PC plant efficiency by more than 10 percentage points.

Flue Gas Contaminant Effect Evaluation:

Flue gas from the PC plant is preprocessed in the Flue Gas Cleanup Subsystem (of the CEPACS plant) prior to its feed to the ECM stacks. The contaminant evaluation addressed possible interactions of the impurities that may be present in flue gas with the ECM, which could result in reduced performance or life. Four main flue gas impurities were considered – sulfur, chlorine, mercury and selenium. The study included thermochemical modeling to predict the possible impurity-membrane interactions, and bench scale experimental work to assess the extent of the interactions, if any. The testing of the ECM cells was conducted by PNNL in their laboratories. FCE provided the button cells required for the tests. The contaminants evaluated included prevalent forms of sulfur, chlorine, mercury and selenium. Effect of these contaminants on ECM cell performance and endurance was studied.

Thermochemical calculations were performed to check the feasibility of contaminants reacting with the ECM cell's carbonate electrolyte system in flue gas (cathode side) and fuel gas (anode side) environments. Specific forms of S, Cl, Hg, and Se which can be present in flue gas were identified. The identified contaminant forms were then assessed as to whether they react with NiO (cathode material) and the electrolyte.

Experimental work to determine if the impurities affect the button cell performance included assessment as a function of concentration and exposure time. Long-term flue gas contaminant exposure tests were performed at a constant current density (i.e. constant CO_2 flux) in the range of 50-160 mA/cm^2, while monitoring the cell voltage. Tests with 0.4 - 1 ppmv SO_2 (in simulated PC plant flue gas used as cathode feed) included a cell operated at a constant CO_2 flux of 176 scc/m^2/s. A steady state was established after ~400 hours of testing, after which no noticeable changes in performance were observed over the remaining 300 hours of testing, even when the SOx concentration was increased from 0.4 to 1.0 ppmv. CO_2 flux remained constant throughout the test. To simulate an upset in the Flue Gas Desulfurization polishing unit of a CEPACS system, higher concentration levels of up to 40 ppm SO_2 were evaluated. While performance losses were observed on introduction of SO_2 at 10 ppm and higher levels, the voltage loss was found to be nearly fully reversible on return to a lower concentration level (1 ppm). Overall, tests showed negligible losses at the system design conditions of ~0.4 ppm SO_2 (in the effluent of the flue gas polisher in a CEPACS system). Button cell tests also included accelerated tests with up to 40 ppm HCl (in ECM cathode feed gas). The purpose of the accelerated tests was to determine the nature of carbonate-chlorine reactions in relatively short test duration. Formation of alkali chlorides was confirmed by post-test analysis. This finding is consistent with thermodynamic predictions. Testing at concentration levels typical of coal power plants was also performed. With 0.2 ppm HCl in flue gas, no measurable cell degradation (at 80 and 160 mA/cm^2) was observed during a 915-hour experiment. No performance losses were observed in tests of up to 1100 hours (at 110 mA/cm^2 and 650°C) due to the presence of 250 ppb Hg(g) in flue gas and up to 750 hours with 250 ppb Hg(g)+0.2 ppm HCl. No performance degradation was observed over a 600-hour test (at 160 mA/cm^2) with 10 ppb SeO_2 in flue gas.

Testing was performed at FCE using a larger 250 cm^2 ECM cell to characterize the ECM's ability to reduce NO_x emissions from the PC plant flue gas. NO_x (as NO) concentration in cathode feed stream (flue gas) was increased from 50 to 216 ppm. At least 70% of the NO_x was removed at 110 mA/cm^2 for the concentration levels tested. Testing also included analysis of the anode exit stream, which confirmed that NO_x was not transferred to the anode side. It was further observed that NO_x had no significant effect on the cell performance at the concentrations tested.

Based on the experimental results, contaminant tolerance levels for the ECM were identified. The contaminant levels expected from the flue gas clean-up (polishing) subsystem were estimated and compared with the ECM tolerance levels. The contaminant evaluation and comparison with CEPACS plant flue gas polishing system output showed that the ECM tolerance levels are well above the contaminant levels expected in the ECM cathode feed gas (treated flue gas).

Membrane Performance Characterization:

A 250 cm^2 ECM cell was fabricated for the membrane performance characterization tests. After completing the qualification testing, system parametric testing was conducted. The testing included operation at various current densities and reactant utilizations (gas feed flow rates). Cathode inlet gas composition from the CEPACS system developed was utilized. Cell performance (measured in volts), CO_2 flux (measured as current) and the % CO_2 transferred were characterized. CO_2 utilization (% CO_2 separated) was determined through mass balances based on the inlet and exit gas compositions. The fuel utilization was varied from 30% to 68%, the CO_2 utilization (%CO_2 separated) was varied from 40% to 92%, and the current density was varied from 29.6 mA/cm^2 to 149.2 mA/cm^2. Cell temperature of 620°C and cathode inlet O_2 concentration of 8.3% (and CO_2 concentrations of 9.9%) were used as the baseline conditions for testing. From these test data, constant flow polarization and constant CO_2 (and fuel) utilization polarization curves were generated for the cell. After the characterization at baseline conditions, cell performance at various cell temperatures and cathode inlet oxygen concentrations was examined. The cell temperature levels studied included 600°C, 620°C and 650°C. The cathode inlet oxygen concentration level was varied from 7% to 10%. Increase in cathode inlet O_2 concentration represents the extent of air supplementation of the flue gas. As the O_2 concentration is enhanced by the air addition, the CO_2 concentration decreases. From the parametric test data collected, the power production and CO_2 flux were calculated. As the current density increases, the power produced increases. As the current density increases, the CO_2 flux (and hence % of cathode gas CO_2 separated for a constant cathode feed flow rate) from the cathode to the anode of the ECM also increases linearly. The results showed that the cell performance increases with the cell operating temperature. The results also showed that the cell performance increased as O_2 concentration increased up to 8.5% level and then began to decrease. This may be due to decreased CO_2 (the other oxidant needed for ECM operation) concentration resulting from increased air supplementation (CO_2 concentration decreases from 11.0 to 8.6% as O_2 concentration is increased from 7.0 to 10.0% in the test). At the baseline system operating conditions of 110 mA/cm^2, 68% fuel utilization, and 92% CO_2 utilization (percent CO_2 separated), a carbon dioxide flux rate of 128 $scc/s/m^2$ and a complete selectivity towards CO_2 transfer from the cathode to the anode were observed.

All this testing was performed up to a fuel utilization of 68%. System simulation results showed improved overall system efficiency at a fuel utilization of 75%. Therefore, an evaluation of the effect of fuel utilization (at 110 mA/cm^2 and 93% CO_2 utilization) on cell performance was performed. The change from 68% to 75% fuel utilization resulted in a 2% decrease in cell voltage. The ECM cell performance was characterized at 75% fuel utilization (93% CO_2 utilization and 8.3% O_2) at a range of current densities. The constant utilization polarization characteristics were generated to guide selection of the ECM operating current density for the CEPACS system.

Balance of Plant Component Technology Gap Evaluation:
BOP equipment and related technology were evaluated to check for commercial availability or need for a custom design. Technical and cost quotations for key equipment were solicited. Technology development, if required, and the extent of development to fill the technology gap were identified as an outcome of the evaluation process. Major components for which the status of technology was examined included Flue Gas Clean-up subsystem, chillers, ECM stack enclosures and interconnections, and CO_2 compressors.

Flue gas clean-up technologies are commercially available and require moderate development for this application. Traditional wet scrubber flue gas desulfurization (FGD) systems are available from many large globally-known OEMs (original equipment manufacturers). Other technologies which are based on dry sorbents also exist. These systems have the potential, based on mass-transfer and reaction kinetics/equilibrium considerations, to achieve ppb levels of SO_2 and HCl after the clean-up. AECOM Corporation has designed a proprietary two-stage FGD system for the CEPACS plant application based on wet scrubber technology which is capable of achieving ppb levels of SO_2 and HCl. FCE also investigated the potential of a single-step, dry sorbent process developed by Hamon to reduce SO_2 and HCl concentrations to ppb levels, as an alternative to the polishing, wet-FGD system designed by AECOM. This technology has been demonstrated on commercial-scale power plants, achieving single-digit ppm SO_2 levels from un-treated flue gas. FCE had discussions with vendors regarding the feasibility and sizing of a commercial-scale dry sorbent process for the CEPACS application.

Key equipment in the CEPACS system for separating CO_2 from other gases present in the CO_2 – rich stream leaving the ECM anodes is the chillers. A survey of vendors indicated that chillers are available from a large number of suppliers for applications requiring chiller capacity up to about 5000 RT (refrigeration tons). Larger capacity chillers (up to 18,000 RT) are needed for the CEPACS system. Therefore, a study to develop the preliminary design and cost estimation for the chiller was contracted to a vendor. A single train (e.g. centralized) chiller system was configured. The design resulted in two stages of chillers at two temperature levels. The first stage chiller design (moderate temperature) resulted in a single 4,400 RT chiller, well-aligned with the size range of the vendor's commercial products (up to 5,000 RT). The second stage of chillers (low temperature), which required 18,000 RT, would have been too large to supply as a single unit. Therefore, the vendor quoted a modular design of five 3,600 RT units based on existing commercial offerings. The estimated chiller costs represented only ~3% of the Total Plant Cost, indicating economic viability for the process.

Vendor contacts were made for compressors (CO_2-rich gas). FCE received multiple proposals from large, globally-known OEMs. The compressors in these proposals comprised state-of-the-art machinery, with a broad range of applications. The performance of the proposed compressor systems was well-aligned with the 85% efficiency assumed in the process model developed by FCE. The efficiencies ranged from 78.5 to 89.5%. When soliciting the technical proposals, all of the OEMs were informed of the hydrogen content in the CO_2-rich stream. The compressors in all of the proposals are suitable for operation at the CEPACS system process conditions. Overall, CO_2 compressors are commercially available and require no development for this application.

As a part of the technology gap analysis, the CEPACS plant piping and ducting including the ECM module interconnections were also evaluated. In the plant layout developed, high (operating) temperature BOP equipment was located proximate to the ECM stack enclosures. Low temperature BOP equipment was centralized in a single area. This approach greatly reduced the lengths of high temperature piping and ducting. The low temperature piping and ducting (including the headers) was designed so that only standard, commercially-off-the-shelf products were used. Carbon steel was used throughout, except where the potential for carbonic acids (due to humid CO_2-containing streams near dew point) required the use of stainless steels. The high temperature piping and ducting were designed in conjunction with the BOP equipment to reduce plant capital and operating costs, and to minimize the heat losses and pressure drops. Vendors familiar with fabricating high temperature ducting for the Power Generation and Chemical Process Industries were identified.

Bench-scale ECM Stack and CEPACS System Test Demonstration:
A bench-scale system was designed and built to conduct demonstration testing to prove the feasibility of using ECM technology for CO_2 capture. The system utilized an ECM stack containing cells with a total electrochemical membrane area of 11.7 m^2. The bench-scale system was implemented via modification of one of FCE's existing test facilities. The facility build consisted of two major components: existing facility modifications and CEPACS skid construction. Construction of the CEPACS skid was completed. All process equipment was fabricated/procured and installed, including: Low Temperature Shift Converter, water separators with level controllers, CO_2 compressor, CO_2 condensing heat exchangers, liquid CO_2 separator, and a chiller capable of supplying -76°F coolant. Fabrication of the ECM stack, including the associated module enclosure hardware was also completed. The test stack included 14 full-area cells. The cell assemblies were obtained from FCE's Torrington, CT commercial fuel cell manufacturing plant.

Bench-scale testing of ECM-based CO_2 capture system was conducted during BP3 of the project. The ECM stack was operated at CEPACS system operating conditions using simulated PC plant flue gas. The total test period of >15,700 h included steady state testing as well as parametric and optimization tests. The stack completed the planned nine months of steady state testing, meeting one of the project milestones. The CO_2 flux of 116 cc/s/m^2 was maintained constant for over 6,500 hours, transferring more than 90% of the CO_2 in the cathode stream to the anode stream. The project technical milestone of verifying CO_2 flux at 100 cc/s/m^2 was achieved, as the CO_2 flux observed in the bench-scale test exceeded the targeted flux. The ECM stack had a nearly constant gross DC power output of nearly 8 kW (>10 kW peak power), and experienced a power degradation rate of 0.05%/1000 h. The power degradation rate was significantly lower than is typically observed for FCE's sub-scale fuel cell stacks operating in power generation mode. The stack transferred more than 120 tonnes (metric tons) of CO_2 from the simulated flue gas to the anode exhaust stream over the total test period, while simultaneously generating more than 110 MWh of gross DC electricity. Parametric testing, including operation at various CO_2 capture percentages and current densities (CO_2 flux), was conducted. The testing showed that the ECM stack is capable of operating at 20% higher CO_2 flux (~140 cc/s/m^2), with a proportional increase in power output. Thermal cycling of the ECM stack was also performed. The stack was cooled from around 600°C (operating temperature) to <100°C. The required three thermal cycles were completed. There was a minimal performance penalty, in terms of DC power produced and no performance penalty in terms of CO_2 flux or current during the thermal cycling evaluation.

Testing of the CO_2 compression/liquefaction skid was carried out in parallel (with the bench-scale ECM stack CO_2 capture tests). The skid was evaluated under simulated ECM stack exhaust conditions, demonstrating the effective liquefaction and separation of high-purity CO_2. Additional testing included optimizing the operation of key system components. Several major advancements occurred in both the physical system hardware and system testing. To effectively evaluate operability at a wide range of operating conditions, system reliability upgrades were completed and any system deficiencies limiting extended duration testing were addressed. Following the system upgrades, a variety of operating conditions were applied to the CO_2 purification skid to characterize the system response, product characteristics and validate the operation. The results in general indicated the ability to produce food grade liquid CO_2

(>99.9% CO_2 and no measureable CO) across all tested operable points. System performance was assessed by gas chromatography (GC) analysis at key locations. By performing mass balance analysis based on the inlet and exit gas flow rates and compositions, it was possible to determine capture rates. Testing was transitioned from system development stage to system prototype maturation, operator training and system optimization for long term data collection processes. Further optimization was made related to the CO_2 capture and purification skid to enable easier operation for extended duration testing and more reliable data collection. A key outcome was the training and validation of several system operators who were not involved with the development of the system hardware. This transition demonstrated the maturity of the CO_2 capture and purification system and its ability to be used as a parametric analysis tool rather than a development prototype. The parametric system data for a wide range of system operating points can be used to validate system models of large scale carbon capture systems and more fully elucidate desirable operating points.

Updated (Final) Technical and Economic Feasibility Study:

The final updated Technical & Economic Feasibility Study was conducted in BP3 of the project to assess the performance and cost of FCE's ECM-based CO_2 capture system. The CEPACS plant utilizing ECM technology was designed to capture and compress >90% of CO_2 from the flue gas of the reference 550 MW (net AC) Pulverized Coal (PC) Subcritical Steam Cycle Plant. Process simulations were updated based on ECM performance realized in bench-scale testing. At the design condition of 128 $scc/s/m^2$, the cell voltage was ~725 mV. Adjusting for cell performance degradation, this beginning-of-life cell voltage would be equivalent to 730 mV, which was used as the basis for the design of the CEPACS system. The updated CEPACS plant contained a total of 1,792 ECM stacks, a reduction of ~10% relative to the BP1 analysis. An equipment list, ECM stacks packaging design, and CEPACS plant layout were developed to facilitate the economic analysis. The economic feasibility study included estimation of CEPACS plant capital cost, cost of electricity (COE) analyses and, estimation of cost per ton of CO_2 captured and cost per ton of CO_2 avoided. The study focused on developing technical and economic comparisons to Cases 9 and 10 related to Subcritical PC plants [26].

The CEPACS system employs 1,792 ECM stacks divided into 8 sections. For large-scale applications of the CEPACS system, grouping the stacks into larger enclosures reduces the overall capital costs by eliminating smaller module enclosures and their associated piping, facilitates the replacement of individual stacks (compared to replacing complete modules) and provides economies of scale that are not possible with 448 separate 1 MW (nominal), 4-stack modules (BP1 stack packaging concept). In each section, 224 ECM stacks are located in an enclosure. The layout of the CEPACS system was designed to minimize capital costs. The "hot" balance-of-plant equipment was de-centralized into 8 separate sections with one section located proximate to each of the 8 ECM enclosures. These "hot" sections comprised such major BOP equipment as the flue gas blower, the flue gas heater, the boiler, the fuel preheater, the air blower, the air preheater, the duct burner and oxidizers, the anode exhaust cooler and the low-temperature shift converter. This modular design minimized the lengths of "hot" piping and the quantities of fittings, significantly reducing capital costs while simplifying the sparing of parts and potentially increasing the capacity factor.

In the process of capturing ≥90% of the CO_2 from the PC plant flue gas, the ECM-based CEPACS system for the 550 MW (net AC) PC plant simultaneously generates 351 MW additional (net AC) power after compensating for the auxiliary power requirements of CO_2 capture and compression. The net electrical efficiency of the CEPACS-equipped PC plant (with CO_2 capture) was estimated to be 38.8% (based on higher heating values of coal and natural gas fuels used by PC plant and CEPACS system, respectively). Cost estimates for all major process equipment and systems were generated from a combination of vendor contacts and

AECOM's historical cost databases. The capital cost for Case 3 (PC + CEPACS-based CO_2 capture) included two components: the CEPACS system cost and the PC plant cost. The PC plant cost estimate was taken directly from the Case 9 estimate as reported in the DOE Baseline Bituminous report [26]. The capital cost estimate for the CEPACS plant was developed by AECOM in 2013/14 dollars. The TOC (total overnight cost) was then de-escalated to 2007 dollars. Cost savings of approximately $20 million were estimated based on redundancy of equipment and utilities when combining the cost estimate for the CEPACS Plant with Case 9 of the DOE report. These estimated savings are primarily attributed to synergies that can be realized by centrally locating redundant equipment systems such as Limestone Feeders/Conveyors/Day Bins, Ball Mills, Dewatering equipment, electrical distribution and plant utilities. These cost savings were deducted from the combined cost estimate. The TOC of the subcritical PC power plant with the CEPACS system for CO_2 capture was estimated to be $2,297/kW. Table ES-2 shows the cost of electricity for Case 3 (PC + CEPACS-based CO_2 capture) as compared to Case 1 (PC w/o CO_2 capture) and Case 2 (PC w/ Amine-based CO_2 capture). The estimated COE for Case 3 is 80.9 mills/kWh. The incremental COE for Case 3 is 36.2%.

Table ES-2 also shows the cost of CO_2 captured calculated for the two CO_2 capture cases. The costs are shown in year 2007 US dollars and per metric ton of CO_2, for consistency with the T&EFS basis. The CEPACS system cost of CO_2 captured ($33.63/tonne) compares favorably with the DOE target of less than $40/tonne (2011 USD). The cost of CO_2 captured for the CEPACS-based process (Case 3) is 27.2% lower than that for the Econamine-based process (Case 2). The cost of CO_2 captured for the amine-based process does not meet the DOE target. The cost of CO_2 avoided for Case 3 is 60.3% lower than that for Case 2, due to the extra power generation of the CEPACS plant which results in significantly lower CO_2 emissions (lb/MWh) than Case 2, at a fixed 90% CO_2 capture rate. Table ES-2 also shows the performance and environmental profile summary for all cases.

Overall, the CEPACS-based CO_2 capture process is more attractive technically and economically than the amine-based process.

Table ES-2. Cost and Performance Summary for All Cases

	Case 1 PC w/o CO_2 capture	PC w/ Amine CO_2 Cap	PC + CEPACS
PERFORMANCE			
CO_2 Capture	0%	90%	92.8%
Gross Power Output (kWe)	582,600	672,700	1,016,064
Auxiliary Power Requirement (kWe)	32,580	122,740	115,239
Net Power Output (kWe)	550,020	549,960	900,825
Coal Flow rate (lb/hr)	437,378	614,994	437,378
Natural Gas Flow rate (lb/hr)	N/A	N/A	124,763
HHV Thermal Input (kWth)	1,495,379	2,102,643	2,321,943
Net Plant HHV Efficiency (%)	36.8%	26.2%	38.8%
Net Plant HHV Heat Rate (Btu/kWh)	9,277	13,046	8,795
Raw Water Withdrawal (gpm/MWnet)	10.7	20.4	8.1
Process Water Discharge (gpm/MWnet)	2.2	4.7	1.9
Raw Water Consumption (gpm/MWnet)	8.5	15.7	6.2
CO_2 Emissions (lb/MMBtu)	204	20	13
CO_2 Emissions (lb/MWhgross)	1,783	217	98
CO_2 Emissions (lb/MWhnet)	1,888	266	111
SO_2 Emissions (lb/MMBtu)	0.0858	0.0017	0.0006
SO_2 Emissions (lb/MWhgross)	0.7515	0.0176	0.0047
NOx Emissions (lb/MMBtu)	0.07	0.07	0.0155
NOx Emissions (lb/MWhgross)	0.613	0.747	0.1205
PM Emissions (lb/MMBtu)	0.013	0.013	0.0042
PM Emissions (lb/MWhgross)	0.114	0.139	0.0329
Hg Emissions (lb/TBtu)	1.143	1.143	0.7360
Hg Emissions (lb/MWhgross)	1.00E-05	1.22E-05	5.74E-06
COST			
Total Plant Cost (2007$/kW)	1,622	2,942	1,881
Total Overnight Cost (2007$/kW)	1,996	3,610	2,297
Bare Erected Cost (2007$/kW)	1,317	2,255	1,480
Home Office Expenses (2007$/kW)	124	213	139
Project Contingency (2007$/kW)	182	369	227
Process Contingency (2007$/kW)	0	105	35
Owner's Costs (2007$/kW)	374	667	416
Total Overnight Cost (2007$ x 1,000)	1,098,124	1,985,432	2,068,815
Total As Spent Capital (2007$/kW)	2,264	4,115	2,618
CO2 TS&M Costs (mills/kWh, 2007$)	0.0	5.8	2.3
Fuel Costs (mills/kWh, 2007$)	15.2	21.3	29.8
Variable Costs (mills/kWh, 2007$)	5.1	9.2	5.5
Fixed Costs (mills/kWh, 2007$)	7.8	13.1	5.1
Capital Costs (mills/kWh, 2007$)	31.2	60.2	38.3
COE (mills/kWh, 2007$)	**59.4**	**109.6**	**80.9**
Incremental COE, Case 1 Basis	0	84.3%	36.2%
LCOE (mills/kWh, 2007$)	75.3	139.0	102.6
Cost of CO_2 Captured ($/ton - 2007)	--	$ 41.89	$ 30.52
Cost of CO_2 Avoided ($/ton - 2007)	--	$ 60.94	$ 24.20
Cost of CO_2 Captured ($/tonne - 2007)	--	$ 46.17	$ 33.63
Cost of CO_2 Avoided ($/tonne - 2007)	--	$ 67.15	$ 26.67

INTRODUCTION AND OBJECTIVES

FuelCell Energy, Inc. (FCE), in collaboration with AECOM Corporation (formerly URS Corporation) and Pacific Northwest National Laboratory (PNNL), has been developing a novel Combined Electric Power and Carbon-dioxide Separation (CEPACS) system [1]. The CEPACS system is based on electrochemical membrane (ECM) technology derived from FCE's carbonate fuel cell products featuring internal (methane steam) reforming and carrying the trade name of Direct FuelCell® (DFC®). The unique chemistry of carbonate fuel cells offers an innovative approach for separation of CO_2 from existing fossil-fuel power plant exhaust streams (flue gases). The ECM-based CEPACS system has the potential to become a transformational CO_2-separation technology by working as two devices in one: it separates the CO_2 from the exhaust of other plants such as an existing coal-fired plant and simultaneously produces clean electric power at high efficiency using a supplementary fuel. Additional information including the operational principle of ECM cell and the process concept for CEPACS system is provided later in this section.

The development effort was carried out under the U.S. Department of Energy (DOE) contract DE-FE0007634. The overall objective of this project was to successfully demonstrate the ability of FCE's ECM-based CEPACS system technology to separate ≥ 90% of the CO_2 from a simulated Pulverized Coal (PC) power plant flue gas stream and to compress the captured CO_2 to a state that can be easily transported for sequestration or beneficial use. In addition, a key objective was to show, through the technical and economic feasibility study and bench scale testing, that the ECM-based CEPACS system is an economical alternative for CO_2 capture in PC power plants, and that it meets DOE's objective related to the incremental cost of electricity (COE) for post-combustion CO_2 capture (no more than 35% increase in COE).

The project was scheduled to be performed in three budget periods (BP). The specific objective for BP1 was to complete the Preliminary Technical and Economic Feasibility Study (PT&EFS). The specific objectives for BP2 were to perform (flue gas) contaminant effect evaluation tests, small area membrane tests using clean simulated flue gas, design a flue gas pretreatment system for processing of the flue gas prior to ECM feed, update the Technical & Economic Feasibility Study (T&EFS) incorporating results of contaminant effect tests and small area membrane tests, and to prepare a test facility for bench scale testing. The specific objectives for BP3 were to perform bench scale testing (parametric and long-duration testing) of 11.7 m^2 ECM CO_2 capture, purification and compression system, and update (as final) Technical and Economic Feasibility Study. In addition, an Environmental Health and Safety evaluation (assessment) of the ECM technology was included. This final technical report presents the progress made under the project. The T&EF study was based on the carbon capture system size suitable for a reference 550 MW PC power plant.

Electrochemical Membrane Technology - Operating Principle

The operating principle of the ECM cell, including the mechanism for transport of CO_2 (by migration of carbonate ions through electrolyte) from the cathode to the anode of the cell, is shown in Figure 0-1, along with the appropriate electrochemical reactions.

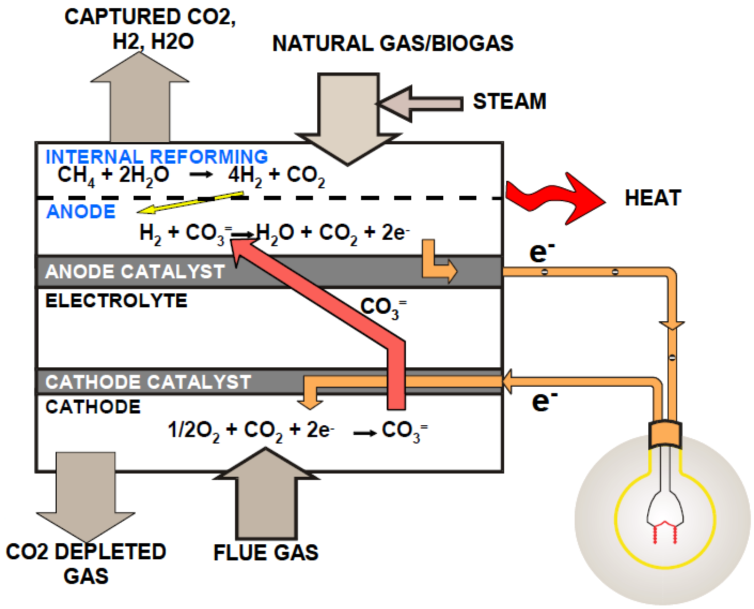


Figure 0-1. Transport of CO_2 in the Electrochemical Membrane Cell: CO_2 is used at cathode as an oxidant and transferred to anode via carbonate ions

The flue gas from an existing power plant, containing CO_2, is fed to the cathode side of the cell. A supplemental fuel is supplied to the anode side of the cell. The ECM technology is compatible with numerous methane-containing fuels such as: coal and biomass derived syngas, natural gas (NG), and biogas (e.g. anaerobic digester gas). Natural gas has been assumed as the supplemental fuel source in this project. Due to the internal reforming capability of the ECM cell, methane in the fuel is converted (steam reformed) into hydrogen according to the following reaction:

$$CH_4 + 2H_2O \rightarrow CO_2 + 4H_2$$

Hydrogen is used as a reactant at the anode. Carbon dioxide and oxygen present in the flue gas are used as reactants at the cathode. The electrochemical reaction at ECM cell cathode Figure 0-1 involves the formation of carbonate ions (CO_3^{2-}) by combination of O_2, CO_2 and two electrons. Carbonate ions produced at the cathode migrate to the anode side via the electrolyte in the cell. At anode, the reaction of carbonate ion with H_2 produces H_2O, CO_2 and two electrons. The internal transport of carbonate ions in ECM cell and the flow of electrons in the external circuit results in power generation as a consequence of the CO_2 separation process. DC power produced is converted to AC power using an inverter.

Overall, the operating mechanism of the ECM cell results in the separation (from flue gas) and transfer of CO_2 into the anode exhaust stream which has a much reduced volumetric flow (resulting in a CO_2-rich stream) compared to the flue gas stream. The CO_2-rich anode exhaust gas is further processed in the ECM-based CEPACS system to purify the CO_2 for sequestration, as described in the CEPACS system process concept portion of this section.

The ECM cell operates at 550 - 650°C and atmospheric pressure. Unlike conventional membrane technologies that rely on pressure (partial pressure) differentials and permeability properties, ECM separates CO_2 at a rate dependent on the electrical current drawn. The flue gas does not need to be pressurized and vacuum operation is not required on the permeate side. Because of fast electrode kinetics, the ECM cell does not require high CO_2 concentration in the flue gas feed. Planar geometry and large gas flow channels of the ECM cell enable processing of large volumetric flow of flue gas feed without significant back pressure. The ECM membrane is fabricated from inexpensive organic materials.

CEPACS System – Process Concept: FCE has developed the Combined Electric Power and Carbon-dioxide Separation system concept (US Patent 7,396,603 B2) as a novel solution for greenhouse gas emission reduction. The CEPACS system employs ECM technology derived from the Company's well established Direct FuelCell® products. A simplified diagram of the CEPACS system concept is shown in Figure 0-2.

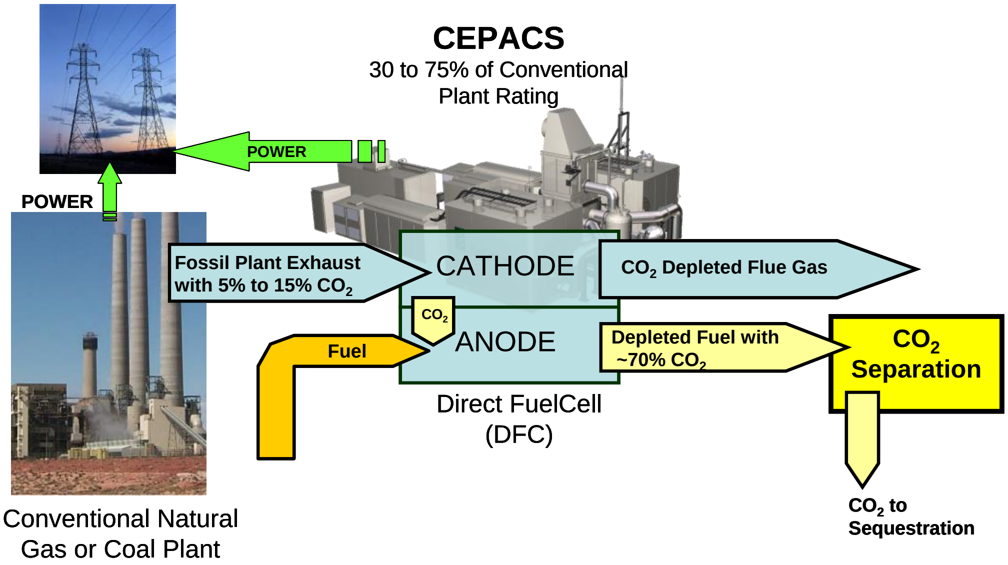


Figure 0-2. CEPACS CO_2 Separation and Power System Concept:
The system can be used with a variety of CO_2-containing greenhouse gases (GHG)

CO_2-containing flue gas from a coal-fired (combustion-based) power plant, such as the exhaust from a PC power plant or other industrial source, is utilized as the oxidant for the ECM cathode. The key feature (as explained in the operating principle of ECM above) is that the ECM utilizes the CO_2 of the flue gas as a reactant for the electrochemical reaction to produce power, while synergistically transferring CO_2 from the flue gas to the anode exhaust stream. A supplementary fuel such as natural gas or biogas (from a digester) is internally reformed in the fuel cell to provide the hydrogen needed to complete the electrochemical power generation cycle.

The CO_2-rich anode exhaust gas is processed (post-processing) in the ECM-based CEPACS system to further concentrate and compress the CO_2 for sequestration. The H_2O (product of ECM anode-side electrochemical reaction) is removed by simple condensation during downstream processing. There is also some unused fuel (mainly H_2) in the CO_2-rich anode exhaust stream. This remaining H_2 is separated by liquefaction of the CO_2. After the water has been condensed out and the H_2 removed, the resulting CO_2-capture stream is ready for compression (pumping of supercritical fluid) and sequestration.

The CO_2-lean cathode exhaust (flue gas after CO_2 removal) is vented to atmosphere after recovering the heat for process use (e. g. preheating of feed streams, steam generation). Water condensed during post-processing of the CO_2-rich anode exhaust stream is used to provide water (steam) needed for internal reforming of CH_4 in supplementary fuel, eliminating the need for external process water. H_2 separated during post-processing of CO_2-rich anode exhaust stream is recycled to provide additional preheat in the system and as part of the supplementary fuel (thereby reducing NG fuel needed). A more detailed process description is presented in Section 1 of this report.

The key difference between the ECM-based CEPACS system technology, and existing membrane and amine scrubbing technologies (competing technologies) is that electric power is produced during the CO_2 separation process in the ECM-based system. This synergistic coupling of the separation of CO_2 from flue gas of an existing plant and the production of clean electric power is unlike any other carbon capture technology, all of which consume energy.

EXPERIMENTAL METHODS

Experimental work was involved in contaminant effect evaluation, performance testing of small-area ECM cells and demonstration testing of bench-scale ECM stack and CEPACS system. The experimental methods used and the testing approach taken are briefly discussed here. Additional experimental details are provided along with relevant test results under 'Results and Discussion' section of the report.

Based on the contaminant effect evaluation, contaminant levels tolerated by ECM cell were determined to define the flue gas clean-up technology needed. Button cells (ECM) provided by FuelCell Energy were used for testing conducted at PNNL. The test stands were built inside the ventilated space equipped with sensor controls appropriate for handling toxic gases. Two furnaces were used, and each could house two button cells. Each cell had a separate gas supply to both of the electrodes using calibrated mass flow controllers. Simulated flue gas was supplied to the cathode side, and simulated reformed natural gas fuel was used on the anode side. Four main flue gas impurities were included in the evaluation – sulfur, chlorine, mercury and selenium. Before introduction of specific impurity/impurities, cells were pre-conditioned at 650°C in clean gas at 0.8-0.9 V for 50-200 hours to obtain a stable baseline, and then were switched to a constant current (constant CO_2 flux) operational mode. After that, the impurity gases from cylinders were added to flue gas near the cell inlet using calibrated electronic flow controllers, downstream of the water bubbler. Mercury was added using a Hg vapor generator. The inlet and exhaust gases were analyzed by Ohio Lumex Hg Analyzer and Jerome Hg analyzers. Both confirmed that outlet mercury concentration was close to inlet Hg concentration.

Long-term exposure tests were performed at a constant current density (constant CO_2 flux) in the range from 50 to 160 mA/cm^2, while monitoring the cell voltage. Arbin Instruments multichannel controllers were used to monitor electrochemical performance of the cells. Electrochemical impedance spectra were obtained at regular time intervals to separate ohmic and electrodic (anodic and cathodic) losses of cells. A Solartron Model 1260/1287 instrument was used to obtain impedance spectra. Each test condition was replicated at least once to confirm reproducibility.

Following termination of electrochemical tests, individual cells were analyzed using scanning electron microscopy and energy dispersive spectroscopy (SEM/EDS). Several cells were mounted into resin and polished for SEM analysis. This provided information on whether alteration phases were formed by electrode interactions with coal gas contaminants and the depth of interaction.

ECM single-cell testing (operation in a CO_2 transfer mode) using simulated flue gas was conducted to verify ECM performance. The testing was performed on a 250 cm^2 (active area) cell in FCE's 7 in x 7 in cell test facility. Cathode inlet gas composition from the CEPACS system simulation was utilized. Cell performance (measured in volts), CO_2 flux (measured as current) and the % of CO_2 transferred were characterized. CO_2 utilization (% CO_2 separated) was determined through mass balances based on the inlet and exit gas compositions.

Parametric testing was performed to evaluate the effects of auxiliary air flow (air supplementation of flue gas), cathode water concentration, current density, CO_2 transfer rate, fuel utilization and cell temperature.

NOx contaminant testing was also conducted using a 250 cm^2 cell. NOx destruction capability of ECM cell was characterized. NO concentration in cathode feed stream was increased from 50 to 216 ppm in five increments for the 100% flow condition. The 100% flow condition corresponds to the oxidant and fuel flow rates required to achieve 92% CO_2 utilization and 68% fuel utilization, respectively at 110 mA/cm^2. Current density was increased from ~46 to 110 mA/cm^2 at each of the six cathode inlet NOx concentration levels. NOx concentration in the cathode exit stream was measured at each operating point. Percent NOx removed from the cathode feed gas was evaluated for the matrix of test conditions. Testing also included analysis of the anode exit stream, which confirmed that NOx was not transferred to the anode side of ECM.

Bench-scale demonstration testing of the ECM system utilized a 14-cell ECM stack with a total area of ~11.7 m^2. The stack was assembled using FCE's full-area cells. An existing 10 kW-scale test facility was modified to conduct the testing. The modified facility accommodated, in addition to the ECM stack module, a CEPACS skid containing equipment for CO_2 (product) stream post-processing. ECM stack conditioning, performance validation and parametric testing were conducted during initial 600 h of operation, before initiating the steady state long-term testing. During the demonstration testing, the CO_2 flux remained constant at 116 cc/s/m^2. Gas chromatography measurements of the ECM stack inlet and outlet streams were performed to estimate and monitor % CO_2 transferred (captured) throughout the testing. DC power output and power degradation rate were monitored during the steady state testing. The parametric testing included operation at various CO_2 capture percentage levels, (anode-side) fuel compositions, stack temperatures and current densities. Thermal cycling was also performed.

RESULTS AND DISCUSSIONS

1.0 PROCESS DESCRIPTION OF ECM-BASED CO_2 CAPTURE SYSTEM

1.1 Overall Process Description

This section describes the overall process for an ECM-based CEPACS system required to process the flue gas from a 550 MW pulverized coal fired Rankine cycle plant. Detailed system modeling was performed using CHEMCAD process simulation software. CHEMCAD-based heat and mass balances were generated. The design for the PC plant is the same as that included in "Case 9" of the DOE - NETL report [2]. A simplified block flow diagram of the CEPACS system is shown in Figure 1-1.

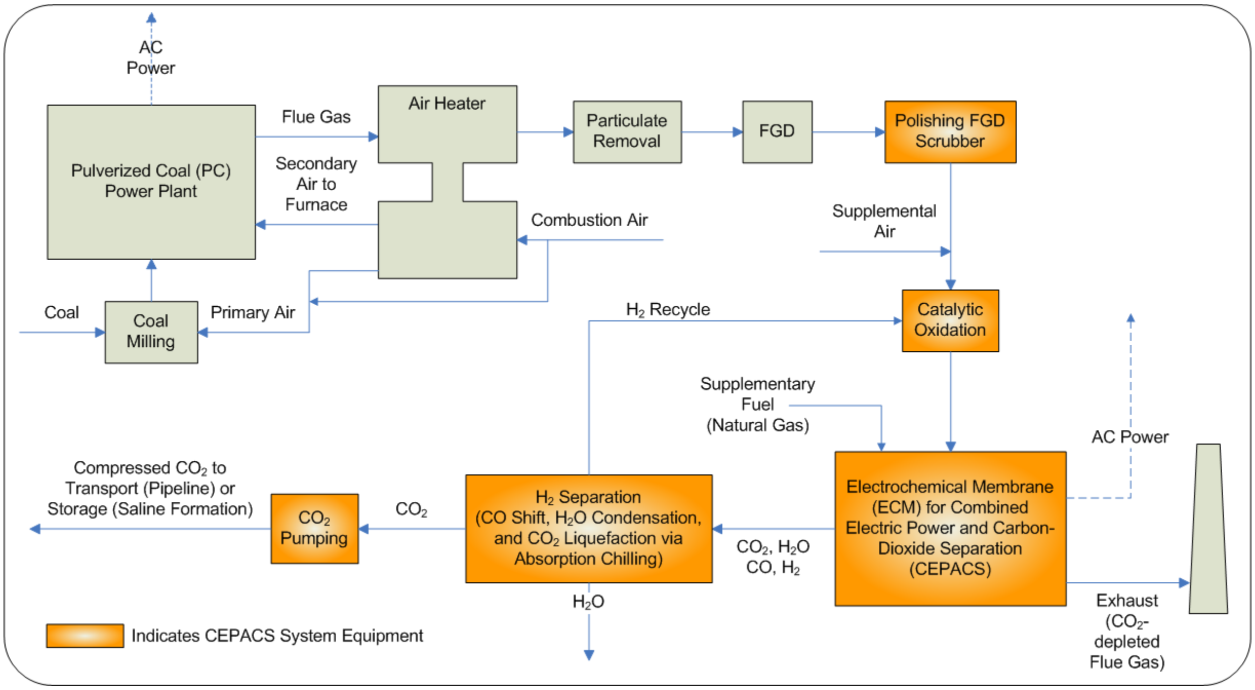


Figure 1-1. Block Flow Diagram of CEPACS System to Process Flue Gas from a 550 MW Pulverized Coal Plant

The flue gas from the PC power plant is routed to the CEPACS system, after the existing flue gas desulfurization (FGD) step. The flue gas is processed in a (secondary) polishing FGD scrubber to remove SO_2 to less than 1 ppmv level. A polishing wet limestone scrubber is expected to exceed the required SO_2 removal efficiency, based on 42 ppmv SO_2 content in the PC plant flue gas. The 1 ppmv SO_2 limit was chosen based on results of Contaminant Effects Evaluation Study, described in Section 3.0. The reasonableness of this tolerance limit is verified by CO_2 capture related experimental work performed by the Chugoku Electric Power Company in Japan (which targeted < 1ppm SO_2 in the feed to molten carbonate fuel cell cathode).

After the polishing FGD scrubber, the flue gas is pre-heated using fuel cell waste heat from the cathode exhaust stream (ECM). Supplemental air is added to the PC power plant flue gas (which is lean in O_2) to boost the O_2 concentration in the stream to a level suitable for ECM operation. The supplemental air is preheated using the heat from anode exhaust (ECM). The mixed flue gas/air stream is directed to a catalytic oxidizer. A hydrogen-rich recycle stream provides the fuel for the oxidizer. The H_2 is combusted in the catalytic oxidizer which increases the flue gas stream temperature to that required by the ECM. The hot stream exiting the oxidizer is then fed to the cathode-side of the ECM. Within the electrochemical membrane modules, CO_2 from the flue gas is transferred from the cathode side to the anode side. For every mole of CO_2 transferred, ½ mole of O_2 is also consumed in the cathode reaction.

The CO_2-depleted stream exiting the ECM cathode provides heat to preheat the natural gas fuel that is fed to the fuel cell anodes and provides heat to generate steam (required for reforming of the natural gas fuel). The cathode exhaust stream also provides heat to the incoming flue gas stream (as mentioned above). The CO_2-depleted cathode exhaust stream is then used to provide heat for a downstream multi-stage absorption chilling system before it is vented to the environment.

Natural gas is supplied as a supplementary fuel to the ECM anodes to drive the simultaneous production of electric power and separation of CO_2. The natural gas is first desulfurized using an (ambient temperature) activated carbon fixed bed. The activated carbon sorbent effectively removes all organic (e.g. mercaptans) and inorganic (e.g. H_2S) sulfur compounds which could damage the ECM anode electrode. The desulfurized natural gas is then mixed with steam and preheated (as mentioned above). The fuel gas enters the ECM module where it is heated to 1100°F (593°C) within the Fuel Superheater (heat exchanger). The Fuel Superheater is an integral component to the ECM stacks located within the module. Methane is internally reformed in ECM stacks, generating H_2, the primary fuel required at the anode. The carbonate ions formed in the cathode reaction migrate to the anode (via electrolyte in ECM cell) and react with H_2 to form water vapor. The electrons liberated in the anode reaction flow to the cathode via an external electrical circuit which completes the power generation cycle. The DC electricity generated by ECM modules (simultaneously while separating CO_2 from flue gas and transferring it to the anode side) is converted to AC power using inverters. The description of the plant electrical system is presented later in this report (Sec 1.2).

The stream leaving the ECM anodes contains the CO_2 transferred from the cathodes, unutilized fuel (primarily H_2 and some CO), and water produced in the fuel cell electrochemical process. This stream is cooled down and directed to a shift converter where the CO is converted to additional CO_2 for capture. The stream leaving the shift converter flows to a condenser where water is recovered for use in the process (includes water feed for steam required - NG reforming mentioned above). The dried stream then enters the first-stage absorption chiller, which utilizes waste heat from the cathode exhaust stream, and is cooled to 35°F (2°C). After additional water separation, the cooled anode exhaust stream is compressed to 250 psia (~1.7 MPa) in multiple stages. In between the compression stages, the temperature is again lowered to 35°F (2°C) using the combination of cooling water and subsequent stages of absorption chilling. After the final stage of compression, a combination of the (hydrogen-rich) recycle stream preheater and the final stage chiller lowers the stream temperature to -47°F (-44°C), at which point the CO_2 is a liquid (supercritical fluid). At this condition, H_2 remains in the gaseous phase and is easily separated from the liquid CO_2. The separation of hydrogen gas from liquid CO_2 occurs in a flash drum. The H_2-rich stream is recycled to the process (after preheating) as mentioned previously. This includes recycle to the catalytic oxidizer (used as fuel for oxidizer as mentioned above) and recycle to the anode fuel feed. The recycle of some H_2 to the ECM anode enables operation of the CEPACS system at high overall fuel utilization (reducing NG fuel use), enhancing the system efficiency. The liquid CO_2 stream is then pumped to 2215 psia (15,062 kPa) and heated (using heat in the anode exhaust stream) to 124°F (51°C) for pipeline transport or sequestration.

The process design features of major subsystems are described in more detail in the following subsection.

1.2 Major Subsystems

Flue Gas Processing Sub-system

Flue Gas Clean-up: The flue gas from the 550 MW PC Power Plant is filtered using a baghouse to remove ash and particulates. The stream is then processed in a wet lime scrubber, referred to as the Flue Gas Desulfurization (FGD) unit, to reduce the SOx (Sulfur) content to 42 ppmV. This is existing process equipment and a part of the PC plant. The flue gas exiting the PC plant requires further sulfur removal to the level of 0.4 ppmV total sulfur to make it suitable for feed to the ECM modules in the CEPACS plant. The additional desulfurization is performed by the Polishing FGD scrubber. A Raw Flue Gas Blower is located upstream of the Polishing FGD to boost the pressure of the flue gas and facilitate its flow through the scrubber unit. The Polishing

FGD is described below.

The polishing FGD is located downstream the primary FGD scrubber. The polishing FGD is a wet scrubber with a design SO_2 removal of 99%. In a wet FGD system, flue gas enters the absorber and is sprayed with wet slurry. SO_2 in the flue gas is absorbed into the sprayed slurry, which is then collected in the reaction tank. In the reaction tank, the absorbed SO_2 combines with an alkaline reagent of limestone to produce gypsum. In order to maintain the pH in the reaction tank, limestone is continuously fed to the tank. The slurry that is collected in the reaction tank is recycled to the spray levels via recycle pumps. A bleed stream of slurry is removed from the reaction tank to control density. The bleed slurry is sent to a dewatering system from where the solids are directed to the same location as the gypsum produced in the primary FGD. The “cleaned” flue gas passes through a mist eliminator, which removes large slurry droplets entrained in the gas stream, before entering the Flue Gas Blower.

The power consumption for the system is estimated to be 4,500 kVA, based on the following assumptions:

- No oxidation air blowers required because SO_2 loading is low
- Limestone feed system not included (assuming limestone is fed from primary FGD system)
- Process water pumps included
- Instrument air compressor included
- Gypsum dewatering not included (assuming gypsum is transferred to same location as the primary FGD)

The water consumption is estimated to be 500 gpm, based on the following assumptions:

- No quenching of oxidation air required
- Limestone feed water included (assuming 30wt% limestone slurry)
- Water consumption for gypsum dewatering not included – assuming water consumption for the system is included in primary FGD system design
- ME Wash Included
- There is no evaporation in the polishing FGD since it is located downstream of primary FGD and the flue gas leaving the primary FGD is saturated with water.

The water discharge from the absorber is assumed to be the same as the water consumption. Since there is no evaporation (due to flue gas saturation), all water added to the absorber is discharged. Most of the water will leave the absorber with the gypsum slurry and the water will therefore be discharged where the gypsum is dewatered.

Pre-heating (Heat Exchangers, Catalytic Oxidizer) and Supplemental Air Introduction: The clean flue gas is heated by passing the stream through the Flue Gas Heater which uses the hot cathode exhaust gas from the ECM modules as the heat source. The Flue Gas Heater is a seam-welded parallel plate heat unit integrated with other heat exchangers (Fuel Preheater, Boiler in the CEPACS plant), forming a series arrangement of three seam-welded parallel plate heat exchangers using a single pass of hot cathode exhaust gas. The cathode exhaust gas exits the Flue Gas Heater at about 492°F (255°C) and is then directed to the ammonia absorption chillers (for use as the heat source). The description for the ammonia chillers is included separately under the CO_2 Post-processing section.

The PC plant flue gas is lean in oxygen. As oxygen from the flue gas is also used as a reactant at ECM cathodes (in the CO_2 capture process), supplemental air is added to the clean flue gas for proper operation of the ECM modules downstream. The supplemental air is preheated to over 900°F (482°C) before it is blended into the hot flue gas. The hot anode exhaust gas from the ECM modules provides the preheat. Some of the heat generated by the Low Temperature Shift Converter (LTSC) is also utilized for preheating of the supplemental air. This is described separately under the CO_2 Post-processing section.

Further preheating of the air-supplemented clean flue gas is carried out in the catalytic oxidizer. Most of the H_2 recycle stream (in the CEPACS plant) is directed to the oxidizer to provide the combustion heat (from exothermic catalytic oxidation) required for the preheating of the flue gas. About 75% of the hydrogen-containing stream remaining after liquefying and separating out CO_2 from the CO_2-rich anode exhaust (ECM) gas is utilized for this purpose. The H_2 recycle stream also contains CO_2 which helps to boost the CO_2 concentration of the flue gas for improved ECM performance. Some flue gas is bypassed around the oxidizer to minimize the size of the parallel oxidizer beds while achieving the maximum exit temperature. The effluent stream from the oxidizer (after the bypass) – clean, air-supplemented, pre-heated flue gas - is now at the required temperature and suitable for feed to the ECM cathodes.

ECM Modules (for CO_2 Separation)

The CEPACS system for separating CO_2 from the flue gas of a 550 MW PC power plant employs 500 ECM modules. A standard module manufactured and used by FCE in a variety of power system applications is shown in Figure 1-2. This module, a major component of the FCE power plant product called DFC1500, is capable of producing 1.2 MW of electricity. Each module includes 4 stacks of carbonate fuel cells (ECM) as shown in Figure 1-3. Each stack has 401 cells.

Figure 1-2. Typical MW-scale Fuel Cell Module (a major component of the FCE power plant product called DFC1500)

Figure 1-3. Fuel Cell Module with Module Enclosure Removed

In conventional power generation applications, the carbonate fuel cells are operated at high power level and ~70% of the CO_2 in the cathode stream is transferred to the anode side. Carbonate fuel cell performance is influenced by the partial pressures of the CO_2 and O_2 (reactants) in the feed stream and utilization levels of the reactants (e.g. % of the CO_2 in the cathode stream that is transferred through the membrane to the anode side) in the cell. For CO_2 separation application, the reactant conditions in the cell result in slightly lower electrochemical performance because of the higher than normal utilization required to achieve greater than 90% CO_2 capture.

The cell performance during operation in a CO_2 transfer mode was determined in laboratory scale testing. Cell voltage was measured as a function of the CO_2 transfer rate or flux defined as cc/s/cm^2. Figure 1-4 shows the resulting performance for three levels of CO_2 utilization (70%, 80% and 90%). The utilization is defined as the percentage of CO_2 transferred (to the anode side) as the stream passes through the cathode of the cell. The CEPACS design with 500 ECM modules removing 90% of the CO_2 in the coal plant flue gas has a CO_2 flux rate of about 0.0135 cc CO_2/s/cm^2 which suggests a cell performance level of about 750 mV based on the lab scale testing presented. Design of the CEPACS system is based on a performance of 790 mV, based on recent improvements in cell performance and operation at lower (than normal) current density (power) level of 110mA/cm^2.

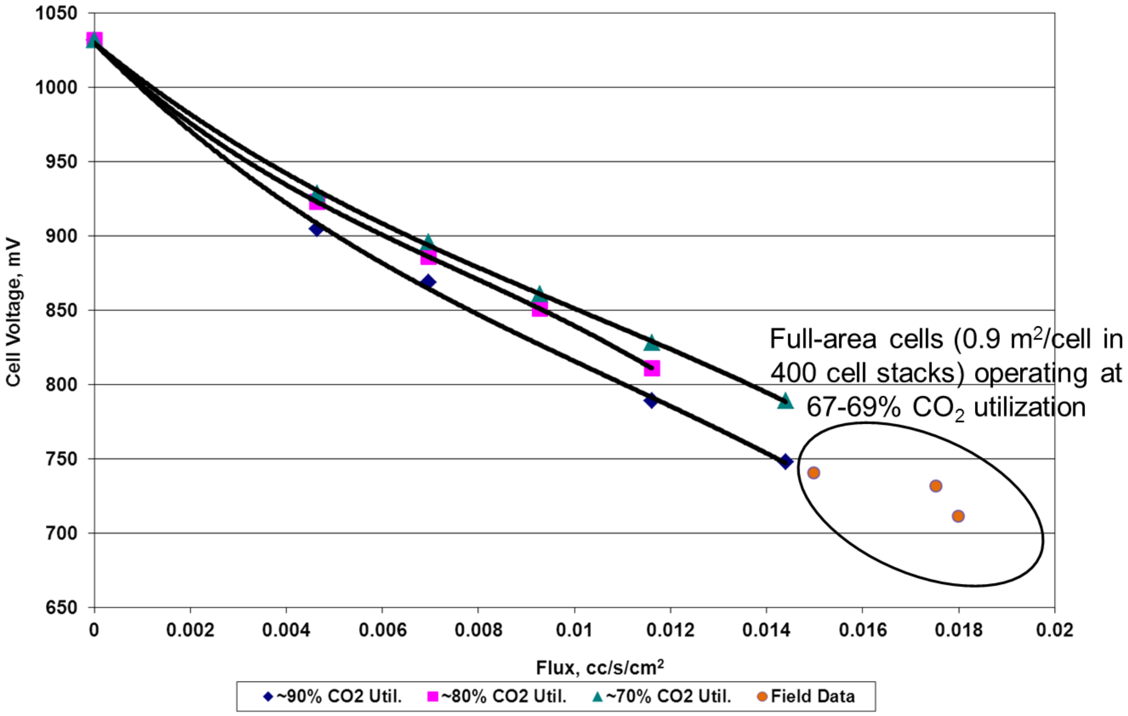


Figure 1-4. Preliminary Cell Performance Based on Subscale Cell Tests

CEPACS Plant ECM Module Layout: Each ECM module is 222 in (5.6 m) long, 161 in (4.1 m) wide and 168 in (4.3 m) tall and weighs approximately 100,000 lbs (45,360 kg). In the CEPACS plant site plan, the 500 ECM modules are arranged in 4 sections as shown in Figure 1-5 . Each section has 125 modules. Roadways are located between the module sections for assembly access. The overall site area required for the 500 modules is 2.49 acres (10,144 m^2).

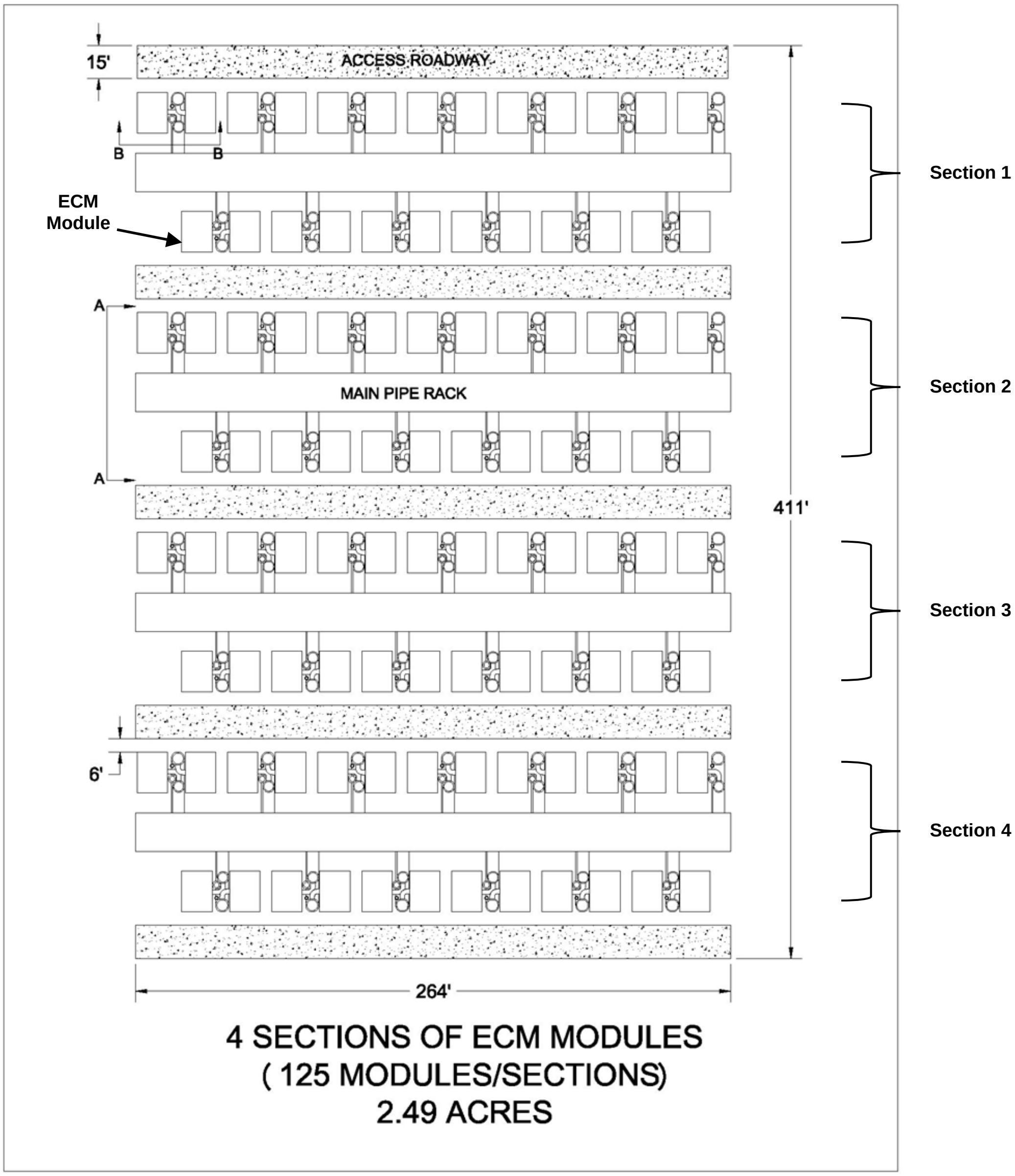
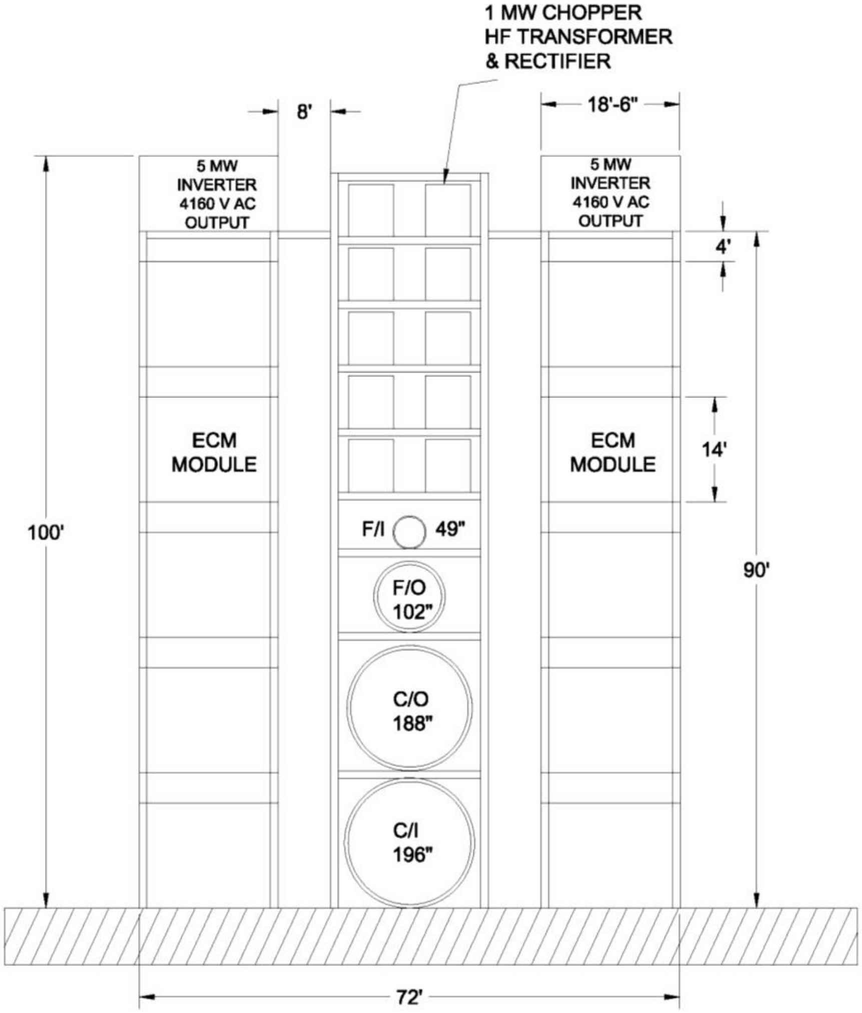


Figure 1-5. Arrangement of 500 ECM Modules in 4 Sections with Access Roads

Figure 1-6 shows an elevation view of a plant section (Section A-A view indicated in Figure 1-5). The modules are arranged in a vertical configuration with five levels (to minimize CEPACS plant footprint) and are supported with a structural steel frame. The 5-module vertical arrays are located on either side of the main gas manifolds that distribute reactants to and collect effluents from each module section. Electrical equipment that converts DC power generated by the ECM modules to 4,160 volt 60 Hz AC is mounted at the top to minimize connection cabling. Figure 1-7 shows Section B-B view (indicated in Figure 1-5). Section B-B view shows the manifolding concept for distributing reactants from the section manifolds to the manifolds that supply the reactants to the 5 modules in a vertical array.



SECTION A-A

Figure 1-6. Section A-A View Showing Arrangement of Modules with Associated Gas Manifolds and Electrical Equipment

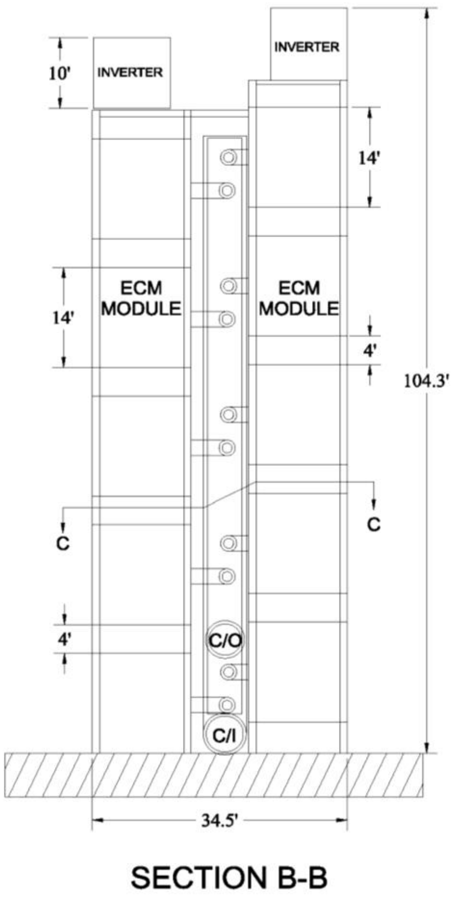


Figure 1-7. Section B-B View Showing Reactant Manifolding for Distribution to 5-Module Vertical Arrays

CO_2 Product Post-processing

Water Gas Shift (Low Temperature Shift Converter) and Water Removal (Cooler, Condenser): The anode gas exiting the ECM modules is CO_2–rich and also contains unused fuel (mostly H_2 and some CO) and product water vapor. It is processed in a Low Temperature Shift Converter (LTSC) to maximize its CO_2 content prior to CO_2 capture steps. Prior to processing in LTSC, the stream is cooled to provide preheat to the supplemental air stream (as mentioned earlier). In addition to the heat exchange in the Anode Exhaust Cooler, the anode gas is cooled further to about 410°F (210°C) by water injection (condensate recycled within the CEPACS plant). The cooled gas is now at the desired temperature to maximize CO shift and is fed to the LTSC where most of the CO is shifted to CO_2, consuming water and generating additional hydrogen (required for its exothermic heat at the oxidizer and recycle to the anode).

The shifted anode gas (from LTSC) is then processed to remove most of the water prior to CO_2 capture steps. The gas is cooled in the Air Preheater to provide pre-heat to the supplemental air. Some of the heat generated by the exothermic shift reaction is utilized here for air pre-heating (as mentioned earlier). The shifted gas is cooled further in the CO_2 (product liquid) Reheater and in the Condenser (using cooling water). This provides substantial condensation and water removal prior to the NH_3–based absorption chilling and pre-chiller compression stages employed for CO_2 capture (described separately below).

Compression, Absorption Chilling and CO_2 Liquefaction: The cooled shifted anode gas stream contains about 79% CO_2. It is then compressed and cooled further in stages to -47°F (-44°C) at 250 psia (~1.7 MPa) to reduce its water content and liquefy CO_2. The intermediate stage features pre-cooling (with cooling water) to 70°F (21°C) and absorption chilling, allowing for water condensation and removal. Compression (to only 250 psia) is carried out in a multi-stage centrifugal compressor. Each compressor train uses electric motor drives.

Two levels of refrigeration are used for the NH_3–based Absorption Chillers. The initial stages provide cooling to 35°F (2°C) and the last stage to -47°F (-44°C). The ECM cathode exhaust gas stream provides all of the process heat needed to drive the low pressure NH_3–based absorption chiller units (desorption stripper heat duties). The existing PC Plant cooling tower water loop is modified to provide 60°F (16°C) water to each compression stage pre-cooler and the Absorption Chiller units (cooling water for absorber and condenser heat loads). The final stage results in a stream that allows for a flash separation of near pure CO_2. The accumulated parasitic power use (by chillers) is relatively small and includes the liquid ammonia and heat transfer fluid pumps.

Product Carbon Dioxide Delivery: High purity CO_2 is flashed as a liquid from the refrigerated and compressed anode gas stream exiting the last chiller at -47°F (-44°C). The off-gas (after the flash operation) containing mostly H_2 and CO_2 is recycled back to the catalytic oxidizer and ECM modules (as mentioned earlier), after preheating in the CO_2 Recycle Heater. The separation of CO_2 (as supercritical fluid) represents over 90% capture of the inlet flue gas CO_2 content. The CO_2 liquid is continuously pumped to 2215 psia (15.3 MPa abs) and heated to the desired temperature of 124°F (51°C) in the CO_2 Reheater.

Supplemental Fuel Processing Sub-system

The CEPACS plant for the CO_2 capture from the 550 MW PC plant is arranged in four sections. The system operates on pipeline natural gas fuel. The natural gas is delivered at a relatively low pressure of about 6 psig (145 kPa abs), as the ECM modules operate at near atmospheric pressure. As it contains ~3 ppmv of total sulfur compounds, the natural gas is desulfurized in a two bed subsystem featuring lead/lag arrangement flexibility. The natural gas flow is valved to the second bed after about 9000 hrs. The sorbent in first bed is replaced and it becomes the second bed in the series flow configuration to guard against preliminary sulfur break-through. Each of the two vessels is about 10 ft (3.05 m) in diameter and 20 ft (6.1 m) high.

The desulfurized natural gas is then mixed with steam at a steam/carbon ratio of about 2.0. Steam is provided from a low pressure boiler that is supplied with water recovered within the CEPACS plant. The humidified natural gas then passes through a fuel preheater which raises its temperature to about 700 °F (371 °C). The fuel preheater and the boiler heat duties are provided by the CO_2-depleted stream leaving the ECM cathodes (as it contains the waste heat generated in the ECM modules). Before distribution to the 125 fuel cell modules in a section of the system, the humidified natural gas is mixed with a hydrogen recycle stream. About 25% of the hydrogen-containing stream remaining after liquefying and separating out CO_2 from the CO_2-rich ECM anode exhaust gas is recycled back to the ECM anodes. The mixed fuel stream to the ECM cell stacks contains ~7% hydrogen. The recycle of hydrogen provides a per-pass fuel utilization in the cells of 75% while achieving an overall fuel utilization of about 80%, which serves to reduce the natural gas fuel consumption.

Water Sub-systems

Process Water (Water Recovery and Recycle within the CEPACS system): This water category refers to high quality water generated and consumed within the CEPACS plant. The anode exhaust gas from the ECM modules contains about 39% water vapor as water is generated as a product of the electrochemical reactions occurring in the modules. A small fraction is consumed in the water gas shift reaction. The remaining water is condensed and separated from the anode stream after the Condenser and the initial Absorption Chiller stages. All condensate is collected in a central tank. Some of this water is used for quenching the anode exhaust stream (from the ECM modules) upstream of LTSC (as mentioned earlier). About 55% (by mass) of the condensate water is pumped to the CEPACS Boiler to generate steam required for ECM module fuel feed. The remaining condensate flow is net excess process water generated by the CEPACS system and is provided to the PC plant for process requirements.

Cooling Water (Addition to PC Plant Cooling Tower): This water category refers to a portion of the PC Plant Cooling tower water which is circulated through the CEPACS heat exchangers and NH_3–based Absorption Chiller units, and does not contact the process gas. This is treated water and normal cooling tower losses are assumed. Supply water is fed at 60°F (15°C) and returned at 80°F (27°C). Two cooling water loops are provided. One loop supplies water to the Condenser and precoolers upstream of the chiller units and requires about 63% of the total cooling water (used for CEPACS). The balance at about 47% of the water is supplied to the Absorption Chiller units (cooling water for absorber and condenser heat loads).

Electrical Sub-systems

The CEPACS system for the 550 MW coal plant includes two electrical subsystems: Power Generation subsystem and Auxiliary Power subsystem. Electrical one-line diagrams were prepared. The Power Generation Subsystem includes 500 ECM modules that generate DC power in the process of separating CO_2 from the coal plant flue gas. The modules are grouped in four sections with 125 modules in each section. Within each section, the modules are arranged in five groups with 25 modules in each group. Each group consists of five 5-module subgroups. Each module has four ECM stacks. The stacks within a module are electrically connected as two parallel lines of two stacks (in series) each. The dc output from each module is converted to 400Hz wave in a square wave chopper. The square wave AC is transformed and rectified. The rectified outputs of the individual modules in the 5-module subgroup are then combined (in parallel) and converted to 4.16 kV in a three phase inverter. The 60 Hz outputs of the inverters (within a 25-module group) are connected in parallel and transformed to 54 kV delivering 30 MVA. The 54 kV outputs from the five 25-module groups are then combined in parallel and transformed to provide the 130 MVA at 345 kV. The outputs from four sections are individually connected to the coal plant switchyard.

The Auxiliary Power Subsystem takes power from the coal plant 345 kV switchyard. An 80 MVA transformer delivers 3-phase 60 Hz power at 54 kV. Auxiliary power for each of the 4 sections in the CEPACS system is transformed to 4.16 KV providing the 20 MVA needed to operate the compressors, blowers, and pumps. Power is transformed to 460 V for smaller pumps, chillers, system controls and auxiliary loads. Power is delivered at 3-phase 60 Hz 277 volts for various single phase loads. Auxiliary power at 4.16 kV is also delivered to the CEPACS flue gas desulfurizer (Polishing FGD) that requires about 4500 kVA.

2.0 PRELIMINARY TECHNICAL AND ECONOMIC FEASIBILITY STUDY

Technical and economic feasibility study (TEFS) of the CEPACS system was conducted to assess the performance and cost of the Electrochemical Membrane (ECM)-based CO_2 capture system. Initially (during BP 1), a preliminary study was performed. This Combined Electric Power (generation) And Carbon dioxide Separation (CEPACS) system employs FCE's Direct Fuel Cell technology which is commercialized for stationary power generation applications. A CEPACS plant utilizing ECM technology was designed for a reference 550 MW (net AC)

Pulverized Coal (PC) Rankine Cycle Plant to capture and compress CO_2 from the PC plant flue gas. The PC plant design specified in "Case 9" of the U.S. Department of Energy - National Energy Technology Laboratory (DOE - NETL) report "*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2" wa*s used as a reference. The system configuration, simulations and analyses were performed using CHEMCAD process simulation software to guide the conceptual design of the CEPACS plant. Table 2-1 presents the flue gas composition and conditions (based on Case 9 of the DOE/NETL report) that were used for the PT&EFS.

Table 2-1. PC Plant Flue Gas Specifications

Flowrate (mass)	5,118,399 lb/h (2,321,705 kg/h)
Pressure	14.7 psia (101.3 kPa abs)
Temperature	135°F (57°C)
Composition (volume %)	
CO_2	13.17
H_2O	17.25
N_2	66.44
O_2	2.34
Ar	0.8
SO_X	42 ppmv
NO_X	74 ppmv

In the CEPACS plant, ECM-separated CO_2-rich stream is cooled, compressed and chilled to liquefy CO_2 which can be easily pressurized for sequestration or beneficial use. Technical information from a leading absorption chiller manufacturer was utilized for system analyses. The performance assessment included estimation of the parasitic power consumption for >90% CO_2 capture and compression, and the efficiency impact on the PC plant. The process flow diagram and a stream table were generated for the CEPACS plant. An equipment list and ECM module layout were prepared to facilitate the economic analysis. The CEPACS plant contained a total of 500 ECM modules.

The economic feasibility study included estimation of CEPACS plant capital cost, cost of electricity (COE) analyses and estimation of cost per ton of CO_2 captured. The following two DOE-NETL documents were used as guidelines for the study/analysis:

- 'Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance', dated April 2011, DOE/NETL-2011/1455 [3]
- 'Scope and Reporting Requirements for NETL System Studies', dated May 2010 [4]

The study was conducted in accordance with the DOE–NETL reports mentioned above, and specifically focused on developing technical and economic comparisons to Cases 9 and 10 of the referenced DOE/NETL Baseline Bituminous report [2]. The capital cost levels and their elements are shown in Figure 2-1. Vendor quotes including the equipment design information and cost were also solicited. Cost estimates for all major process equipment and subsystems were generated from vendor contacts and AECOM's historical cost databases. The cost estimate included the cost of consumables and preliminary manufacturing costs for unique process equipment items.

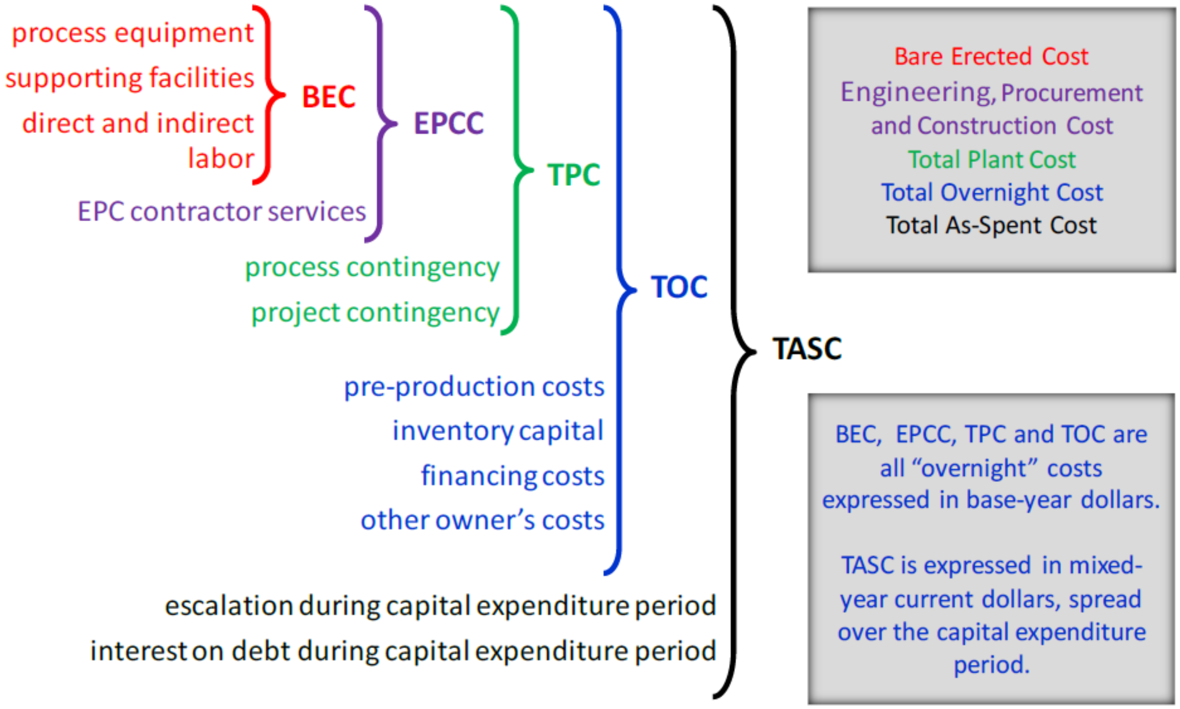


Figure 2-1. Capital Cost Levels and Their Elements

The economic analysis assumptions are listed in Table 2-2. The economic assumptions were taken from three sources: 1) NETL Cost Estimation Methodology [3], 2) Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity – Revision 2, DOE – NETL report [2], and 3) FuelCell Energy's Statement of Project Objectives for this project. Primary goal was to develop the cost estimates on an equal basis to those reported in the referenced Baseline Bituminous Coal report, allowing for a fair and accurate comparison between the cases.

Table 2-2. Economic Analysis Assumptions

Parameter	Value	Basis (see notes)
Levelized-Cost of cooling water, \$/1000 gal	0.12	1
Levelized-Cost of Process Make-up Water, \$/1000 lbs	0.07	1
Levelized Cost of Solid-Waste Disposal, \$/ton	17.87	1
Levelized-Cost of Toxic-Waste Disposal, \$/ton	89.36	1
Levelized Cost of CO_2 Transportation, Storage, & Monitoring, \$/ton CO_2	4.05	1
Plant On-Stream (Capacity) Factor, days/y (%)	310.25 (85%)	1, 2, 3
Natural Gas Price, \$/MMBtu	6.55	3
Plant Location	Generic site, Mid-west	1, 2, 3
Dollar-Year Reporting Basis	2007	1, 2, 3
Total Fixed O&M Levelized-Costs, \$/calendar day	995	1
Maintenance-Material Costs (% of initial equip cost)	2	1, 2, 3
Capital Expenditure Period	5 years	2, 3
Capital Charge Factor, %/yr	12.4 (IOU, high-risk)	2, 3
Levelization Factor (IOU @ 12% IRROE)	1.268	2, 3
TASC/TOC Multiplier	1.140 (IOU, high-risk)	2, 3

1) FCE Project SOPO

2) NETL Cost Estimation Methodology (DOE/NETL-2011/1455)

3) Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity – Revision 2, 2010 (DOE/NETL-2010/1397)

System reconfiguration to improve thermal integration, and optimization studies with respect to the operating parameters to increase efficiency and reduce cost (plant capital cost and cost of electricity), were conducted. ECM parameters such as current density, fuel utilization, cell performance (voltage) level and the number of ECM modules/plant were varied for the optimization. While the ECM-based CEPACS system for the 550 MW PC plant captures 90% of CO_2 from the flue gas, it generates additional (net AC) power after compensating for the auxiliary power requirements of CO_2 capture and compression. The net electrical efficiency of the PC plant equipped with a CEPACS system (for CO_2 capture) was estimated to be 39.8% (based on higher heating values of coal and natural gas fuels used by PC plant and CEPACS plant, respectively). COE for the CEPACS-equipped (for carbon capture) PC plant was estimated to be 80.5 mills/kWh (2007 USDollars). The incremental COE (compared to PC plant with no carbon capture) was 35.5%.

System cases studied are listed in Table 2-3. In addition to the baseline ECM case above (Case 3 in Table 2-3), two alternative cases were studied. One case (Case 4) evaluated the performance and cost of a natural gas combined cycle (NGCC) plant integrated with the 550 MW PC plant, with the combined plant featuring Econamine-based CO_2 capture. This provided a comparison with the ECM-based CEPACS case as in this study the CEPACS system uses pipeline natural gas as a supplementary fuel. The other alternative case (Case 5) considered a smaller PC plant integrated with a CEPACS plant for >90% CO_2 capture and compression. The total net AC output of the combined plant was set to be 550 MW.

Table 2-3. Case Summary

Case ID	DOE/NETL-2010/1397 Case ID [2]	Unit Cycle	MW gross	MW net	Boiler Technology	Sulfur Removal/ Recovery	CO_2 Separation
1	9	PC	583	550	Subcritical PC	Wet Flue Gas Desulfurization (FGD)	--
2	10	PC	675	550	Subcritical PC	Wet FGD	Amine Absorber
3	--	PC + ECM	1106	991	Subcritical PC	Wet FGD	ECM-based CEPACS
4	10 + 14	PC + NGCC	1186	1024	Subcritical PC + HRSG	Wet FGD	Amine Absorbers
5	--	PC + ECM	553	496	Subcritical PC	Wet FGD	ECM-based CEPACS

Table 2-4 shows the cost, performance, and environmental profile summary for all cases.

Table 2-4. Cost and Performance Summary for All Cases

	Case 1 PC	Case 2 PC w/ Amine CO_2 cap	Case 3 PC + CEPACS	Case 4 PC + NGCC w/ Amine CO_2 cap	Case 5 ½(PC + CEPACS)
PERFORMANCE					
CO_2 Capture	0%	90%	90%	90%	90%
Gross Power Output (kWe)	582,600	672,700	1,106,140	1,183,700	553,070
Auxiliary Power Requirement (kWe)	32,580	122,740	114,968	160,170	57,484
Net Power Output (kWe)	550,020	549,960	991,172	1,023,530	495,586
Coal Flow rate (lb/hr)	437,378	614,994	437,378	614,994	218,689
Natural Gas Flow rate (lb/hr)	N/A	N/A	150,756	167,333	75,378
HHV Thermal Input (kWth)	1,495,379	2,102,643	2,489,282	3,208,455	1,244,641
Net Plant HHV Efficiency (%)	36.80%	26.20%	39.82%	31.9%	39.82%
Net Plant HHV Heat Rate (Btu/kWh)	9,277	13,046	9,041	10,696	9,041
Raw Water Withdrawal (gpm/MWnet)	10.7	20.4	8.9	14.8	8.9
Process Water Discharge (gpm/MWnet)	2.2	4.7	2.4	3.5	2.4
Raw Water Consumption (gpm/MWnet)	8.5	15.7	7.0	11.4	7.0
CO_2 Emissions (lb/MMBtu)	204	20	12	18	12
CO_2 Emissions (lb/MWhgross)	1,783	217	94	161	94
CO_2 Emissions (lb/MWhnet)	1,888	266	104	186.4	104
SO_2 Emissions (lb/MMBtu)	0.0858	0.0017	0.0006	0.0012	0.0006
SO_2 Emissions (lb/MWhgross)	0.7515	0.0176	0.0044	0.0122	0.0044
NOx Emissions (lb/MMBtu)	0.07	0.07	0.0144	0.0511	0.0144
NOx Emissions (lb/MWhgross)	0.613	0.747	0.1107	0.5374	0.1107
PM Emissions (lb/MMBtu)	0.013	0.013	0.0039	0.0090	0.0039
PM Emissions (lb/MWhgross)	0.114	0.139	0.0302	0.0963	0.0302
Hg Emissions (lb/TBtu)	1.143	1.143	0.6865	0.7922	0.6865
Hg Emissions (lb/MWhgross)	1.00E-05	1.22E-05	5.27E-06	8.46E-06	5.27E-06
COST					
Total Plant Cost (2007$/kW)	1,622	2,942	1,803	1,971	2,338
Total Overnight Cost (2007$/kW)	1,996	3,610	2,201	2,450	2,848
Bare Erected Cost	1,317	2,255	1,419	1,492	1,845
Home Office Expenses	124	213	133	140	173
Project Contingency	182	369	217	255	279
Process Contingency	0	105	34	84	41
Owner's Costs	374	667	398	479	510
Total Overnight Cost (2007$ x 1,000)	1,098,124	1,985,432	2,181,683	2,508,760	1,411,584
Total As Spent Capital (2007$/kW)	2,264	4,115	2,509	2,793	3,247
CO_2 TS&M Costs	0.0	5.8	2.2	4.6	2.2
Fuel Costs	15.2	21.3	30.7	35.6	30.7
Variable Costs	5	9.2	6.4	6.1	7.7
Fixed Costs	7.8	13.1	4.6	9.6	6.6
Capital Costs	31.2	60.2	36.7	40.8	47.4
COE (mills/kWh, 2007$)	**59.4**	**109.6**	**80.5**	**96.8**	**94.5**
Incremental COE, Case 1 Basis		84.3%	35.5%	62.9%	59.1%
LCOE (mills/kWh, 2007$)	75.3	139.0	102.0	122.7	119.8

3.0 FLUE GAS CONTAMINANT EFFECT EVALUATION

This section presents the results of the Contaminant (effect) Evaluation performed for the Electrochemical Membrane (ECM) cells. Flue gas from the PC plant is preprocessed in the Flue Gas Cleanup Subsystem (of the CEPACS plant) prior to its feed to the ECM stacks.

The contaminant evaluation addressed possible interactions of the impurities that may be present in flue gas with ECM cell, which could result in reduced performance or life. Four main flue gas impurities were considered – sulfur, chlorine, mercury and selenium. Specific forms of S, Cl, Hg, and Se which can be present in flue gas were identified. The study included thermochemical modeling to predict the possible impurity-membrane interactions, and bench scale experimental work to assess the extent of the interactions, if any. The testing of the ECM cells was conducted by Pacific Northwest National Laboratory (PNNL) in their laboratories. FCE provided the button cells required for the tests. The contaminants evaluated included prevalent forms of sulfur, chlorine, mercury and selenium. Effect of these contaminants on ECM cell performance and endurance was studied. The effort also included post-test analyses to aid in understanding and supporting the electrochemical cell test results.

3.1 Thermodynamic Assessment of Impurity Interactions

Thermochemical calculations were performed to check the feasibility of contaminants reacting with the ECM cell's carbonate electrolyte system in flue gas (cathode side) and fuel gas (anode side) environments. As a complement to experimental studies, these calculations helped assess potential reactions of impurities with ECM cell components. Impurities considered were sulfur, chlorine, fluorine, mercury, and selenium. Figure 3-1 summarizes impurity interactions that were assessed, including direct reaction with the cathode, interactions with the molten carbonate electrolyte in equilibrium with flue and fuel gases, and reactions that would allow the impurity to reach the anode. Thermochemical calculations were performed using FactSage 6.4 and associated databases.

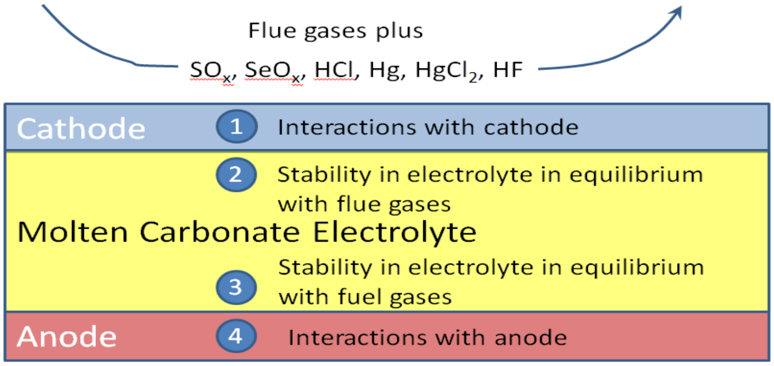


Figure 3-1. Impurity Interactions with ECM Cell

Sulfur: In flue gases, sulfur is present primarily as a mixture of SO_2 and SO_3, the latter of which is favored at lower temperatures. Sulfur in flue gases could potentially degrade the performance of nickel oxide in the cathode by reaction to form nickel sulfate:

$$NiO(s) + SO_3(g) \rightarrow NiSO_4(s) \quad (1)$$

Critical (minimum) concentrations of gaseous sulfur species required to form $NiSO_4(s)$ were estimated as a function of temperature. At 650°C, ~500 ppm total sulfur is projected to be necessary to form the solid alteration phase, approximately two orders of magnitude higher than expected sulfur concentrations in flue gas after cleanup. As such, formation of solid reaction products is considered to be highly unlikely.

Displacement of alkali carbonate by sulfate through reactions with oxides of sulfur proceeds according to the following reaction:

$$\{M\}_2CO_3(l) + SO_3(g) \rightarrow \{M\}_2SO_4(l) + CO_2(g) \quad (2)$$

where {M} represents the metallic cation of the carbonate salt.

The Gibbs free energy of reaction for reaction (2) was estimated to perform gas-liquid equilibrium calculations. Minimum SO_3 partial pressure (in flue gas) necessary to form $\{M\}_2SO_4$ was estimated to be 5×10^{-13} atm, lower than that present in flue gas, even following cleanup. Therefore, it is reasonable to conclude that some SOx may be captured by the molten carbonate electrolyte.

In addition to reaction (2), SO_2 and SO_3 may be captured directly in electrochemical reactions, following:

$$SO_2(g) + \tfrac{1}{2} O_2(g) + 2 e^- \rightarrow SO_3^{2-}(l) \qquad (3), \text{ and}$$

$$SO_3(g) + \tfrac{1}{2} O_2(g) + 2 e^- \rightarrow SO_4^{2-}(l) \qquad (4).$$

Alkali sulfates are expected to be highly miscible in the molten carbonate electrolyte. The ternary phase diagram (based on calculations) for the $\{M\}_2CO_3$-$\{M\}_2SO_4$,system has an extensive region of liquid stability. As such, any sulfates that are formed from SO_x capture are expected to remain in the liquid phase rather than to form a separate solid phase.

Equilibrium concentrations of sulfur-containing gas species were calculated for carbonate electrolyte and $\{M\}_2SO_4$ mixtures at 650°C. The sum of all gaseous sulfur species remained below 10 parts per trillion even if one quarter of the carbonates in the liquid phase had been replaced with alkali sulfates. These results show that alkali sulfates are highly stable when dissolved in molten carbonate electrolytes in equilibrium with flue gases.

While sulfur oxides are expected to be readily captured by the molten carbonate electrolyte at the cathode side, sulfur components are unstable in the melt at the anode side, and are expected to be lost (expelled) as H_2S.

The primary means for sulfur compounds to be expelled from the molten carbonate electrolyte is as $H_2S(g)$ on the anode side. Both electrochemical and non-electrochemical routes are possible. Electrochemical elimination of sulfate at the anode can be expressed as:

$$SO_4^{2-}(l) + 5 H_2(g) \rightarrow H_2S(g) + 4 H_2O(g) + 2e- \qquad (5).$$

The overall electrochemical reaction for sulfur uptake at the cathode side and sulfur elimination on the anode side is:

$$SO_3(g) + 0.5 O_2(g) + 5 H_2(g) \rightarrow H_2S(g) + 4 H_2O(g) \qquad (6).$$

Elimination of sulfate at the anode side could also occur without the electrochemical steps:

$$\{M\}_2SO_4(l) + 4 H_2(g) + CO_2(g) \rightarrow H_2S(g) + \{M\}_2CO_3(l) + 3 H_2O(g) \qquad (7).$$

The presence of $H_2S(g)$ in fuel gases is known to result in diminished MCFC performance [5-15]. The formation of nickel sulfide phases, particularly $Ni_3S_2(s)$, is expected to require much higher $H_2S(g)$ concentrations than are likely to be encountered. The predominance diagram given in Figure 3-2 shows that approximately 1000 ppm $H_2S(g)$ is needed to form $Ni_3S_2(s)$ at 650°C.

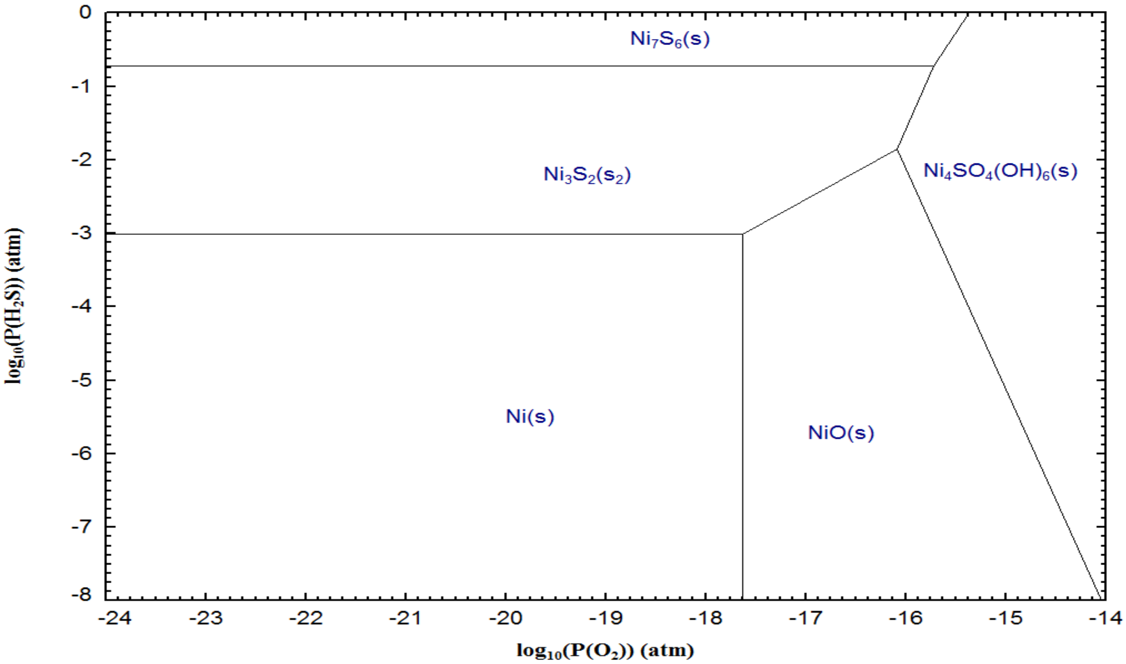


Figure 3-2. Predominance Diagram for Nickel as a Function of H_2S(g) and O_2(g) Partial Pressures at 650°C: Formation of bulk nickel sulfide phases is not expected at anticipated conditions

The following is a summary of calculations on exposure to SO_x in flue gases:

(1) Formation of $NiSO_4$(s) at the cathode would require more than 500 ppm SO_x(g), and thus is not expected.

(2) There is a large driving force for uptake of SO_x(g) by the molten carbonate electrolyte on the cathode side. Alkali sulfates are highly soluble in the molten carbonate, when equilibrated with flue gases, and can readily migrate within the electrolyte from cathode (flue gas side) to the anode (fuel side).

(3) Sulfates are unstable in molten carbonate electrolytes when equilibrated with fuel gases. Sulfates are expected to be reduced at the anode and eliminated from the electrolyte in the form of H_2S(g), where it may interact with the anode. However, formation of bulk nickel sulfide is expected to require approximately 1000 ppm at 650°C, and thus is not expected.

Chlorine: Nearly all of the chloride salts will be converted to HCl(g) in coal-derived combustion gases [16, 17]. The distributions of chloride species in flue and fuel gases are given in Figure 3-3, assuming an initial concentration of 1 ppm HCl(g).

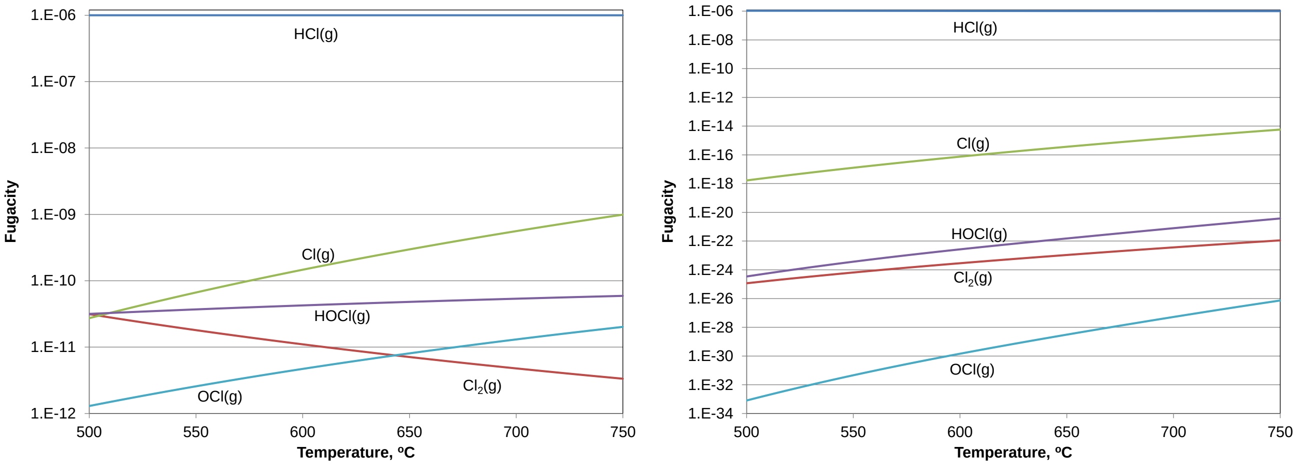


Figure 3-3. Chlorine Speciation in Flue (left) and Fuel (right) Gases versus Temperature: Species HCl(g) predominates in both

Reaction of HCl(g) with either NiO(s) in the cathode or Ni(s) in the anode to form bulk chlorides is not expected under ECM operating conditions. A predominance diagram for the Ni-Cl-H-O system is shown in Figure 3-4. More than 100,000 ppm HCl(g) is required to form bulk $NiCl_2$(s) under either oxidizing or reducing conditions.

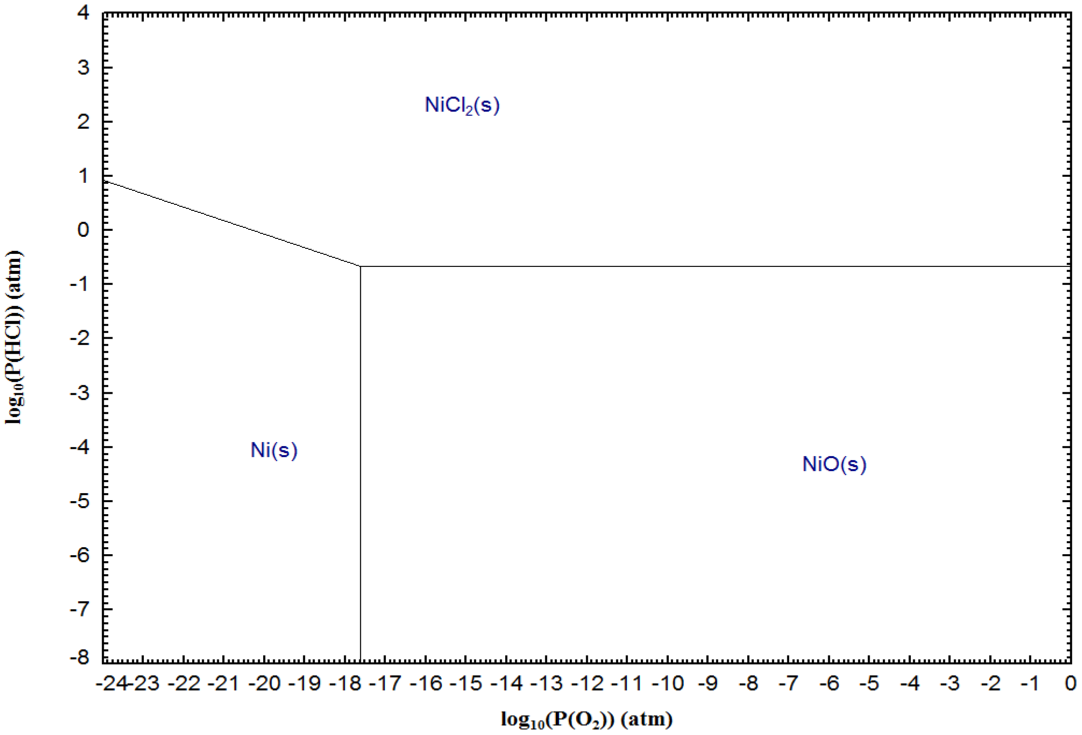


Figure 3-4. Predominance Diagram for Ni-Cl-H-O System at 650°C: No new chloride-containing bulk phases with nickel or nickel oxide are expected under realistic conditions

As with oxides of sulfur, there is a strong thermodynamic driving force for the displacement of carbonates in the electrolyte by chlorides. Chlorine may be captured by the electrolyte following:

$$2\ HCl(g) + \{M\}_2CO_3(l) \rightarrow 2\ \{M\}Cl(l) + CO_2(g) + H_2O(g) \qquad (8)$$

The free energy of reaction for reaction (8) was estimated. There are no obvious electrochemical reactions involving chlorine oxidation/reduction that could result in enhanced chloride uptake.

A phase diagram for the $\{M\}_2CO_3$-$\{M\}Cl$ system at 650°C was generated, which showed a broad compositional range in which a single liquid phase is present. Solid chlorides are expected to form for total alkali chloride concentrations >0.4 ppm.

Alkali chloride is the most volatile species from carbonate electrolyte containing dissolved alkali chloride. Fugacities of chloride-containing gases in equilibrium with molten carbonate electrolyte at 650°C were estimated for both the flue gas and fuel gas environments. For long-term exposure of the electrolyte to 1 ppm HCl(g), the expected mole fraction of alkali chlorides is ~0.05, leading to the conclusion that most of alkali chlorides escape ECM membrane in the vapor phase. Figure 3-5 shows an estimate of the quantity of carbonate electrolyte expected to remain after extended exposure to flue gases containing varying concentrations of HCl(g). However, for low HCl concentrations (ppb levels) expected after the flue gas polishing subsystem (in the CEPACS plant), no significant loss of the carbonate electrolyte is expected in 10-15 years of ECM operation.

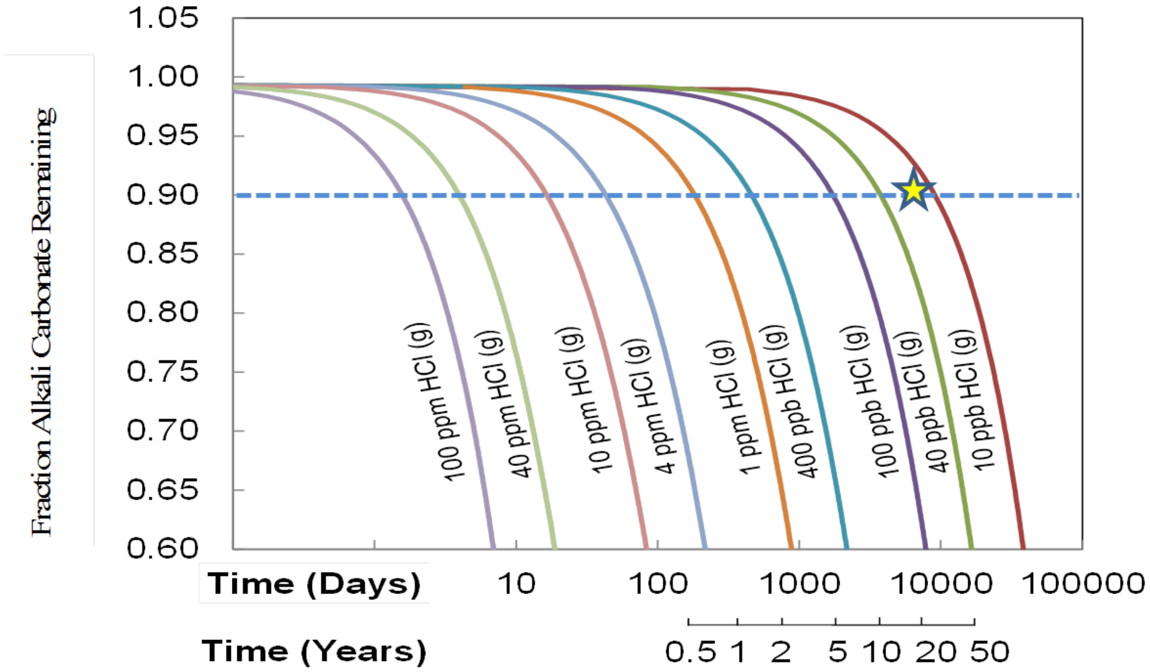


Figure 3-5. Estimated Molten Carbonate Quantity Remaining Following Exposure to Flue Gas Containing HCl(g)

The following is a summary of calculations on ECM exposure to HCl(g) in flue gases:

(1) Bulk nickel chloride is not expected to form from the reaction of either nickel metal or nickel oxide with anticipated HCl(g).

(2) Chlorides are thermodynamically favored to displace carbonate from the molten salt electrolyte.

(3) Chlorides are eliminated from the electrolyte through volatilization of alkali chlorides, which could lead to some electrolyte loss, depending on exposure time and HCl(g) concentration. At expected HCl concentration levels after flue gas cleanup, no significant electrolyte loss is expected in 10-15 years of operation.

Fluorine: Most of the fluoride is volatilized when coal is combusted, the predominant form of which is HF(g) [18]. The emissions standards for fluorides from coal-fired power plants are typically <1 ppm [19].

Fluorides show little tendency to react with either metallic nickel in the anode or NiO in the cathode to form bulk phases. A predominance diagram for the Ni-H-F-O system given in Figure 3-6 shows that Ni and NiO are stable to >100,000 ppm HF(g), much higher than the expected fluoride content even without cleanup.

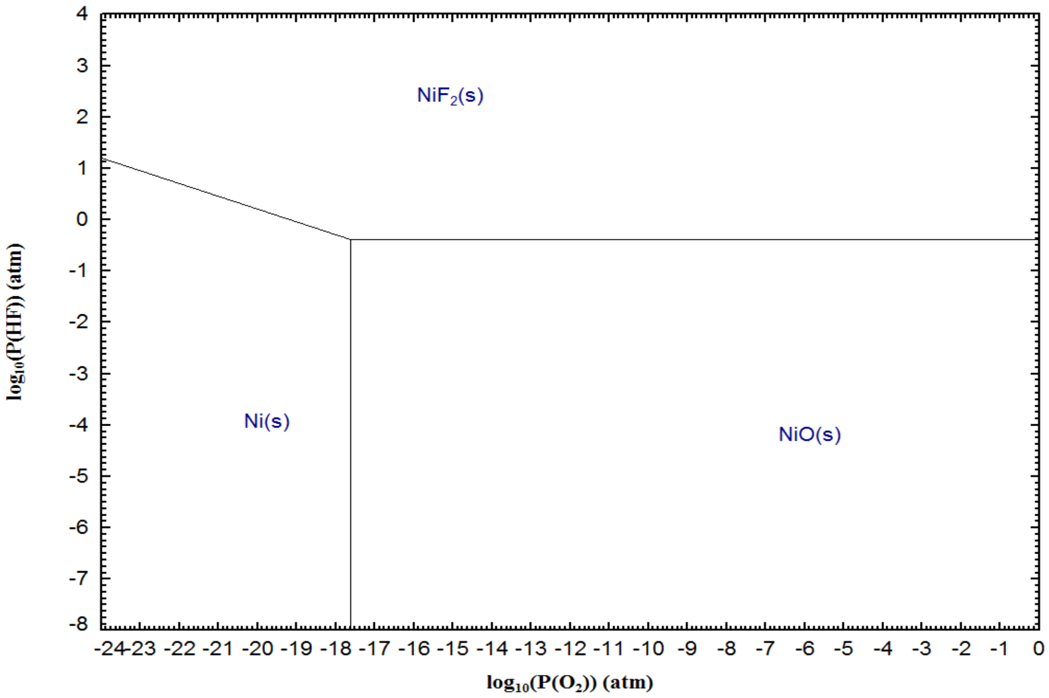


Figure 3-6. Predominance Diagram for Ni-H-F-O System at 650°C.

Displacement of carbonates by fluorides is favored, following:

$$2HF(g) + \{M\}_2CO_3(l) \rightarrow 2\{M\}F(l) + CO_2(g) + H_2O(g) \qquad (9)$$

The free energy of reaction for reaction (9) at 650°C was estimated. There are no viable electrochemical pathways for fluoride uptake, because the oxidation state of fluorine in the gas state is the same as that in the melt. The ternary phase diagram (based on calculations) for the $\{M\}F$-$\{M\}_2CO_3$ systems at 650°C revealed broad compositional ranges of a single liquid phase. Alkali fluorides are highly soluble in the molten carbonate electrolyte. The alkali fluoride solids are not expected to form below {M}F mole fraction of ~0.8.

Concentrations of fluoride-containing gases in flue gas in equilibrium with a molten carbonate electrolyte that contains varying concentrations of alkali fluorides were estimated. The predominant fluorine-containing species is HF(g), whereas alkali fluoride gas species are found in concentrations two orders of magnitude smaller than HF(g). This is important because, unlike interactions with chlorides, the fluorides should not lead to loss of electrolyte through volatilization.

Equilibration of the molten carbonate electrolyte in flue gases containing 1 ppm HF(g) will lead to an {M}F mole fraction of ~0.13. Similarly, long-term exposure to flue gas with an HF(g) concentration of 0.1 ppm would lead to an {M}F mole fraction of ~0.02. Negligible impact of fluoride at these concentrations is expected.

The following is a summary of calculations on ECM exposure to HF(g) in flue gases:

(1) Bulk nickel fluoride is not expected to form from the reaction of either nickel metal or nickel oxide with anticipated HF(g).

(2) Unlike chlorides, fluorides are not expected to result in alkali loss from the electrolyte.

Mercury: Interactions between mercury and other impurities in flue gases, especially chlorine, can enhance volatility of mercury [20, 21]. Figure 3-7 provides speciation for mercury in flue gases with and without 40 ppm HCl(g) as a function of temperature. In the absence of HCl(g), the predominant specie is Hg(g), followed by HgO(g) with approximately an order of magnitude less. With HCl(g) introduction, $HgCl_2$(g) is found, which predominates at low temperature. Figure 3-8 showes mercury speciation at 650°C in flue gases as a function of HCl(g) fugacity, and shows $HgCl_2$(g) to predominate for HCl(g) fugacities greater than ~250 ppm. Because flue gases are expected to be cleaned to significantly less than 1 ppm HCl(g), no synergistic impact of Hg(g) by presence of HCl(g) is anticipated. The sensitivity of mercury to the presence of SOx(g) in flue gases was also assessed. No impact is expected.

No reactions of mercury vapor in either flue or fuel gases were identified in thermodynamic assesements using available databases. Similarly, uptake of mercury compounds by the molten carbonate electrolyte was not predicted. Thermodynamic solution parameters that would enable an assessment of miscibility in molten carbonates are not available. As such, experimental studies were necessary to show whether the presence of mercury with and without other impurities results in a performance loss.

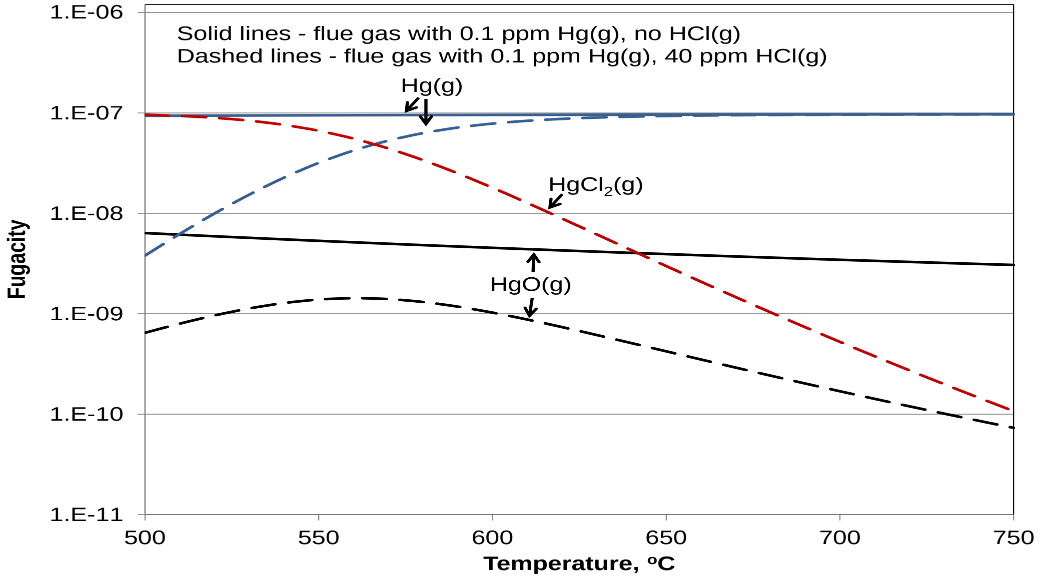


Figure 3-7. Mercury Speciation in Flue Gas (with & without HCl) as a function of Temperature for Initial Hg(g) Concentration of 0.1 ppm

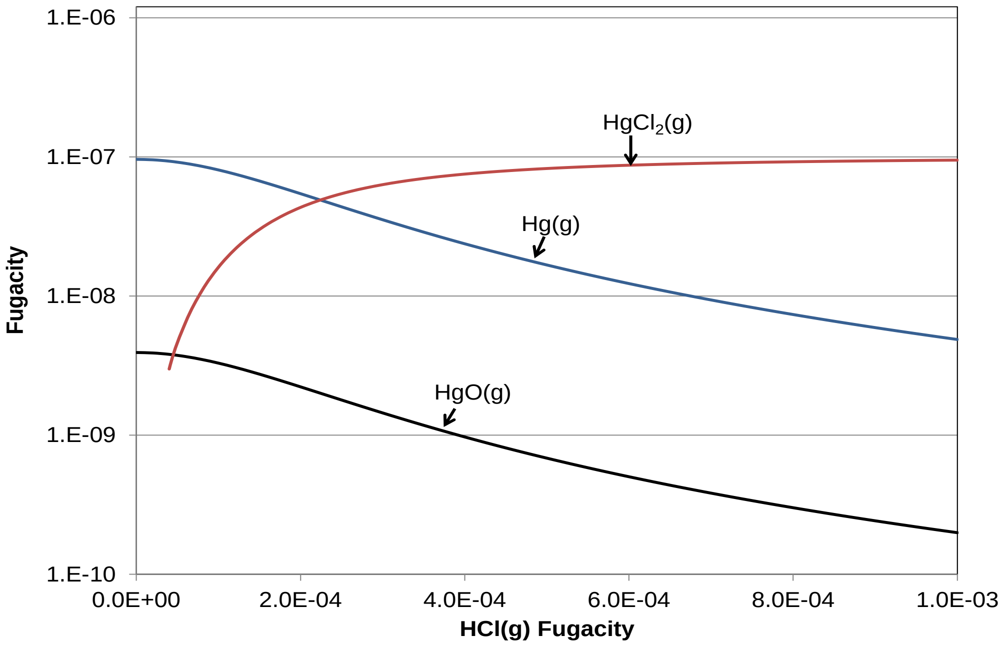


Figure 3-8. Mercury Speciation in Flue Gas at 650°C as a Function of HCl Fugacity, for Initial Hg(g) Concentration of 0.1 ppm: $HgCl_2(g)$ is expected to predominate for HCl(g) concentrations > ~250 ppm, far higher than anticipated concentrations in flue gas following cleanup

The following is a summary of calculations on ECM exposure to Hg(g) in flue gases:

- In the absence of other impurities, the predominant form of mercury in flue and fuel gases is Hg(g).
- The presence of HCl(g) leads to the formation of $HgCl_2(g)$, favored by low temperatures and high HCl(g) concentrations. At 650°C, Hg(g) will predominate for HCl(g) concentrations less than 250 ppm.
- No reactions of mercury with either NiO in the cathode or Ni in the anode are expected.

Selenium: In flue gases, selenium is expected to be converted to SeO_2(g) [20,22,23]. Figure 3-9 shows speciation for 1 ppm Se in fuel and flue gas, indicating H_2Se(g) to predominate in reducing conditions and SeO_2(g) to predominate in oxidizing conditions.

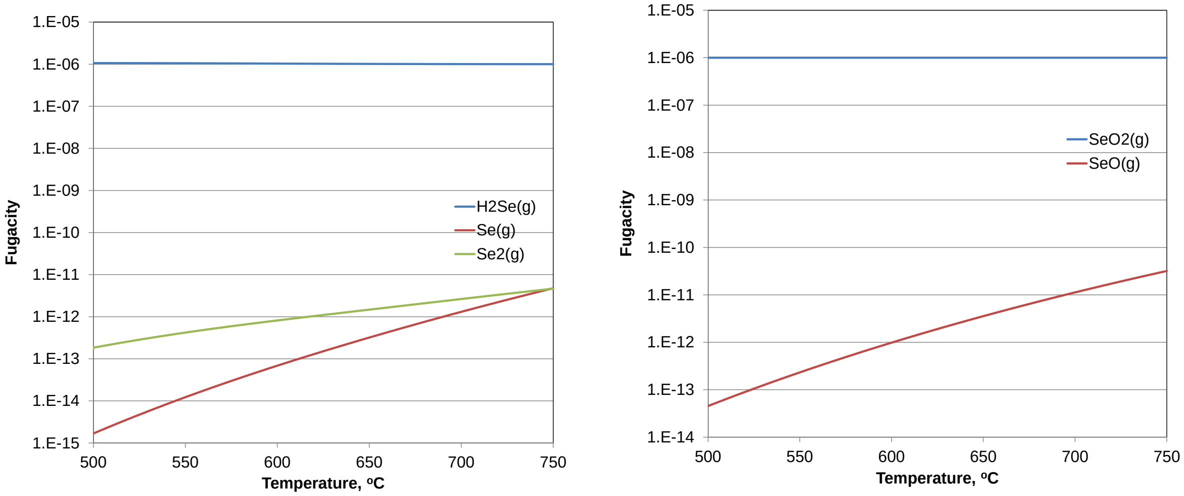


Figure 3-9. Predominant Forms of Selenium in (gas phase of) Fuel Gas (left) and Flue Gas (right) for Selenium Concentration of 1 ppm

Substantially higher than realistic concentrations of selenium in flue gases are needed to result in the formation of bulk reaction products such as $NiSeO_3$(s). A predominance diagram for the Ni-H-O-Se system was generated which showed that nearly 1% selenium would be required to form such products. Similar to sulfur oxides, selenium dioxide may displace alkali carbonates. Data were not available for liquid states of $\{M\}_2SeO_3$ or $\{M\}_2SeO_4$. Therefore, approximations and assumptions were, utilized for the analysis. Also, calculations on the solubility and stability of selenites and selenates could not be performed due to a lack of thermodynamic solution parameters.

Analogous to sulfates, selenates and selenites may be thermodynamically unstable in fuel gases, and may be expelled from the electrolyte as H_2Se(g). Parameters for the liquid state of $\{M\}_2SeO_3$ and $\{M\}_2SeO_4$ were not available. If selenium may be captured in the molten carbonate electrolyte and transported to/reduced at the anode, in a manner similar to that described for sulfur, interaction with the anode is possible. Figure 3-10 shows that the solid phase Ni_3Se_2(s) may be formed in fuel gases containing >10 ppm H_2Se(g) at low fuel utilization, and as little as 0.5 ppm H_2Se(g) at high fuel utilization. A similar analysis for SOFC anodes in coal gas has been given previously [24, 25].

The following is a summary of analysis on ECM exposure to SeO_2 in flue gases:

- SeO_2(g) in flue gases is not expected to react with nickel oxide in the cathode to form new bulk phases.
- Capture of SeO_2(g) by the electrolyte is likely, through displacement of carbonate salts. Though thermodynamic solution parameters are not available for oxidized selenium liquid species, the likely form is as alkali selenites and/or selenates. These species will probably migrate from the cathode (flue gas side) to the anode (fuel side).
- $\{M\}_2SeO_x$ species are thermodynamically unstable in fuel gases, and are expected to decompose to form H_2Se(g) in reactions analogous to sulfur uptake and exsolution.
- Nickel selenide bulk reaction products are not expected at realistic selenium concentrations in flue gas.

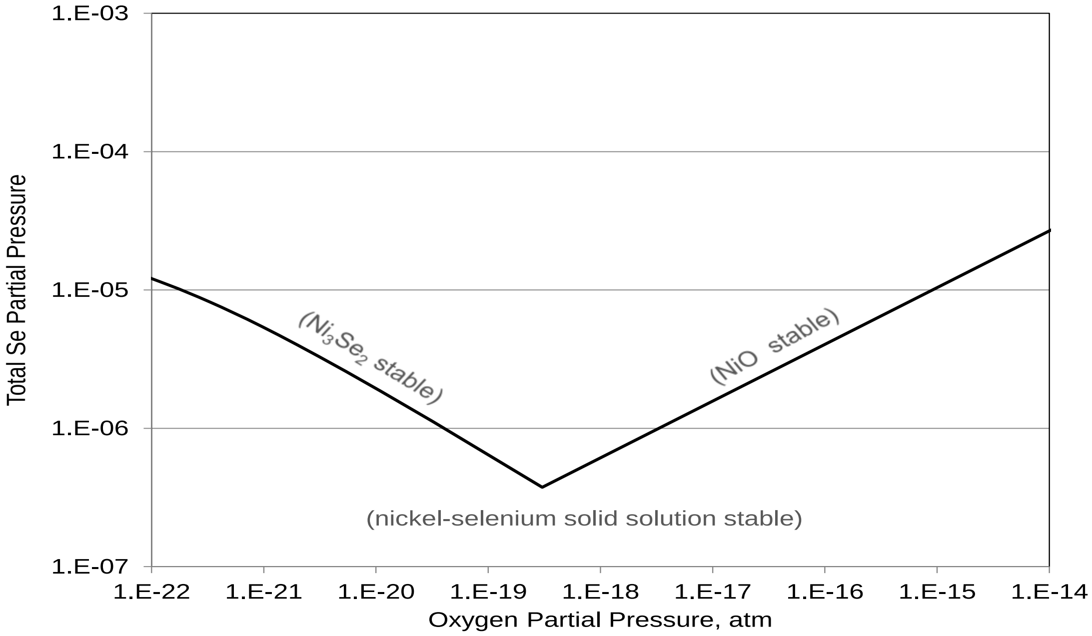


Figure 3-10. Total Selenium Partial Pressure in Fuel Gas Required to Convert Nickel Anode to Ni_3Se_2(s) or NiO: As fuel utilization increases, lower concentration of selenium is needed to form reaction products

3.2 ECM Cell Testing with Feed Gas Containing Impurities

Experimental work to determine if the impurities affect the button cell performance included assessment as a function of concentration and exposure time. Long-term flue gas contaminant exposure tests were performed at a constant current density (i.e. constant CO_2 flux) in the range of 50-160 mA/cm^2, while monitoring the cell voltage. Each test condition was replicated at least once to confirm reproducibility.

Sulfur: A button cell test was conducted to evaluate ECM cell performance with simulated flue gas containing SO_x. Concentration levels of 0.4 and 1 ppm SO_x were included in the evaluation. Figure 3-11 shows the test results. After establishing a baseline performance at 0.9 V, the cell was operated at a constant current density of 125 mA/cm^2, initially on clean gas and then with 0.4 ppm SO_x added to simulated flue gas. In this discussion, sulfur is indicated as SO_x which represents an equilibrated mixture of SO_2 and SO_3, present in flue gas. The cell operated with typical voltage gain in the initial stages of operation (up to ~180 hours) after SO_x was introduced. Continued operation of the cell showed the cell establishing its expected steady state cell voltage, with no significant loss observed during the following 245 hours of exposure to 0.4 ppm SO_x.

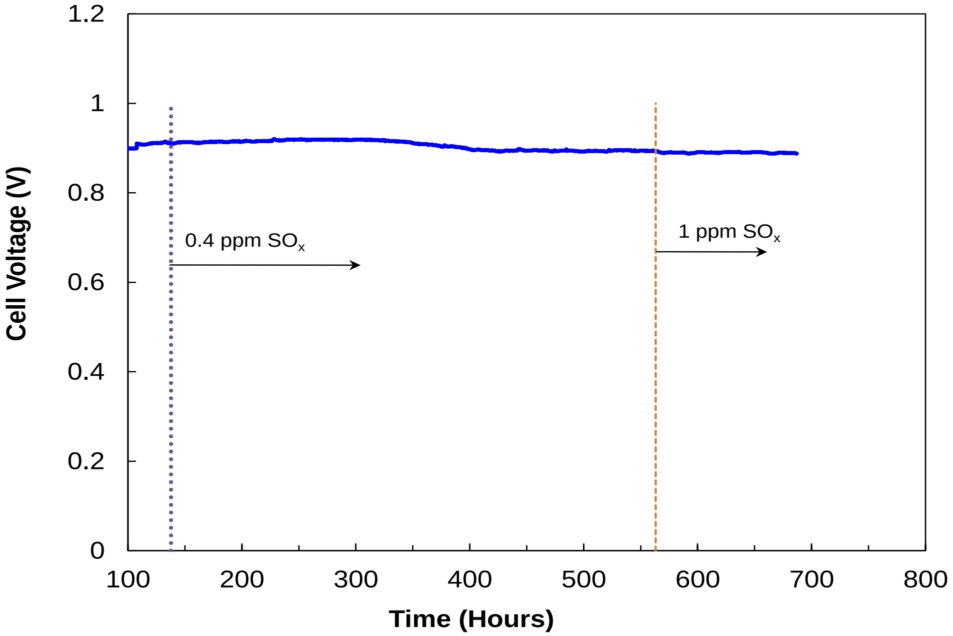


Figure 3-11. ECM Cell Performance at Constant Current Density of 125 mA/cm^2 (650°C) During Operation on Simulated Flue Gas Containing 0.4 ppm and 1 ppm SO_x

The total test time with 0.4 ppm SO_x was 425 hours. After that the sulfur concentration was increased to 1 ppm and the test was continued for the next 175 hours. No noticeable changes in the performance were seen. This test was reproduced several times and the results were similar. In another test, no significant loss was observed in presence of 0.5 ppm SO_x at 160 mA/cm^2, and the increase in SO_x concentration to 1 ppm had minimal effect on the cell voltage. Long-term cell performance data with 1 ppm SO_x in simulated flue gas were collected for up to 610 hours twice and the results of one of the tests are illustrated in Figure 3-12. Results of these tests can be summarized as follows:

1) No noticeable performance loss occurred when 0.4 ppm or 0.5 ppm SO_x was added to the flue gas. The long-time exposure to SO_x (<0.5 ppm) concluded no further changes in the performance of the ECM membrane as compared to normal behavior without presence of sulfur in the flue gas.
2) Similar response was observed in presence of 1 ppm SO_x. Stable operation and constant CO_2 flux was confirmed in 500-600 hour tests.
3) The overall behavior of the ECM performance stability in presence of sulfate formation in electrolyte is explained by the impedance spectroscopy tests. The proposed mechanism could be described as the following:
    a. sulfur reacts with the electrolyte to form $\{M\}_2SO_4$;
    b. $\{M\}_2SO_4$ diffuses to the anode side, where it is reduced to H_2S, per reaction (7). The presence of H_2S in the anode gas was confirmed by mass spectrometry.
    c. H_2S may reversibly adsorb on the Ni anode with minimal change in the rate of the electrochemical reactions.
4) No long-term ohmic resistance increases were observed on operating cells indicating no permanent sulfur accumulation in the electrolyte.

5) The experimental results were reproducible and were consistent with the thermochemical modeling results.

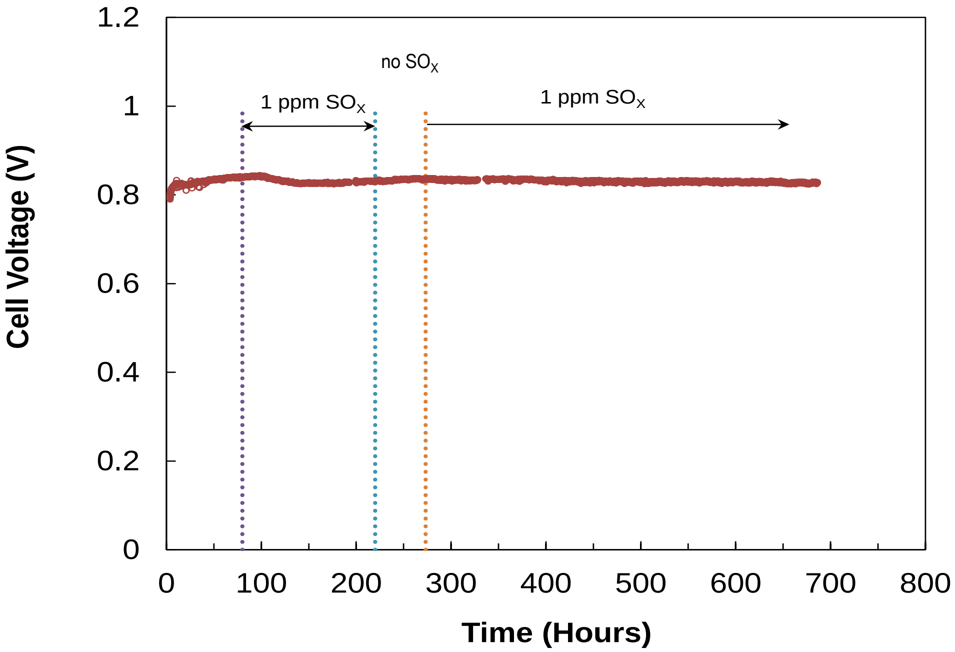


Figure 3-12. ECM Cell Performance at Constant Current Density of 152 mA/cm^2 (650°C) During Operation on Simulated Flue Gas Containing 1 ppm SO_x

Tests were also conducted to study ECM cell response to exposure to higher sulfur concentrations (10-40 ppm SO_x) which may be caused by temporary operational upsets in the flue gas desulfurization system. Figure 3-13 shows the results of exposure to 10 ppm SO_x for various lengths of time. After establishing a steady baseline, 1 ppm SO_x was added to simulated flue gas. Only a minor voltage loss of < 1% was observed and a new steady state was established. Then the sulfur concentration was increased from 1 ppm to 10 ppm for 1 minute and reduced back to 1 ppm, as shown in the upper part of the plot. No change in the cell voltage was observed. The sulfur concentration was then increased to 10 ppm for 10 min and decreased back to 1 ppm. Again, no voltage loss was detected. The procedure was repeated with 30 and 60 minutes of exposures to 10 ppm SO_x that resulted in a 3 and 10 mV voltage losses, respectively. The voltage loss was fully reversible, once the sulfur concentration was reduced to 1 ppm. When changing the sulfur concentration, the total gas flow rate was maintained constant by adjusting the nitrogen flow rate. The testing was continued to evaluate exposures (to 10 ppm SO_x) longer than one hour. When the cell was exposed to 10 ppm SO_x for nearly 7 hours (400 minutes), the voltage loss was 140 mV (~17%). Cell performance fully recovered when the SO_x concentration was decreased to 1 ppm. During nearly 17 hours (1000 minutes) of exposure to 10 ppm SO_x, the performance loss was the same as during the 7-hour exposure, indicating that the maximum Ni anode poisoning (maximum Ni coverage) by sulfur had likely already occurred. During a shorter exposure of about 3 hours (170 minutes), only partial poisoning occurred followed by a quick recovery. A slight performance improvement from the least active state was noticed in the 17-hour exposure test and attributed to sulfur participating in the charge transfer reaction. With high concentrations of sulfates in the electrolyte matrix, SO_4^{2-} can participate in electrochemical reactions:

$$SO_4^{2-} + 5H_2 \rightarrow H_2S + 4H_2O + 2e^- \qquad (10)$$

Again, the cell fully recovered when SO_x concentration was decreased to 1 ppm, although it took longer to recover. Thus, the cell recovers fast when partial sulfur coverage is reached and slower after the full coverage. This test showed that both performance losses due to high SO_x concentrations and the recovery time are dependent on the exposure time. SO_x at 10 ppm level caused minimal effect during short (≤ 1 hour) exposures and led to performance losses (<20%) during longer exposures (3-17 hours). The losses were fully reversible. Figure 3-13 also illustrates that the cell fully recovers from even long exposures to 10 ppm SO_x.

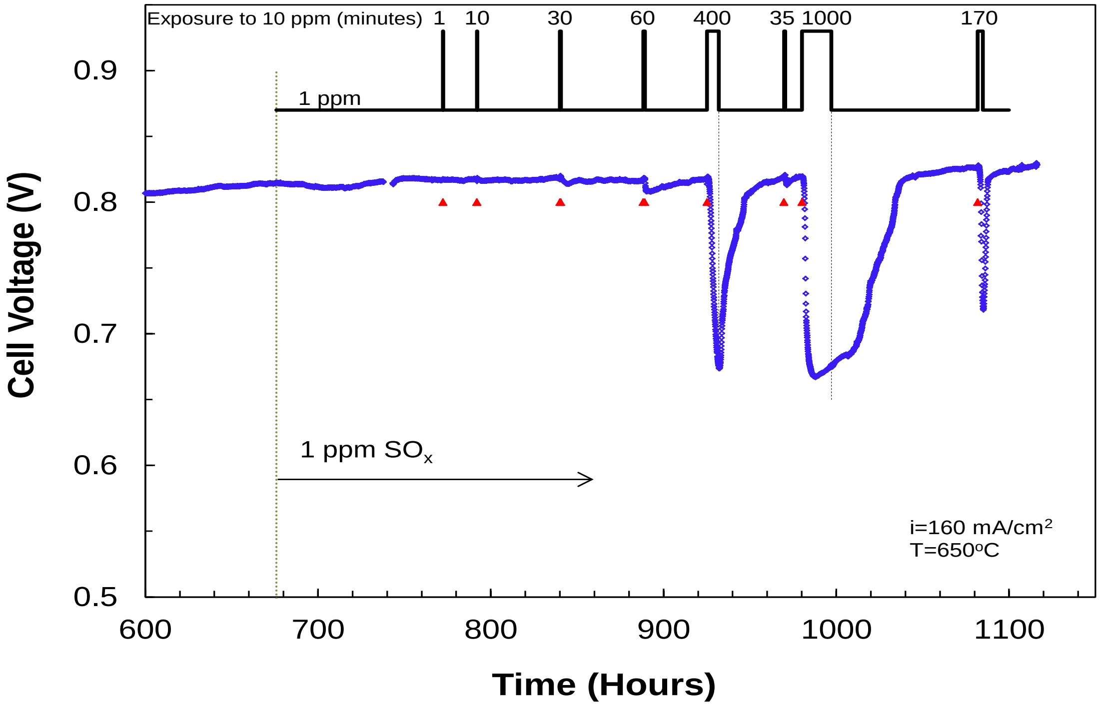


Figure 3-13. ECM Cell Response during SO_x Concentration Excursions to 10 ppm (times when SO_x was increased are marked by red triangles, times when it was reduced back to 1 ppm after prolonged exposures are marked by dashed lines)

Post-test SEM/EDS analysis detected no sulfur or sulfur compounds on either side of the matrix or the electrodes after 600-hour exposure to 1 ppm SO_x in flue gas. No sulfur compounds were found also on the cell exposed to higher SO_x concentrations, up to 40 ppm. This is consistent with the performance data (showing no permanent damage to the cell, which could be associated with the presence of new phases) and with the thermochemical modeling results.

Chlorine: When 0.2 ppm chlorine (15-20 times higher than expected downstream of the flue gas desulfurization system) was added to simulated flue gas, no changes in performance were observed during the testing duration of over 916 hours at both low and higher currents, as shown in Figure 3-14. ECM cell voltage is a function of the operating current density. Cell voltage decrease observed with increase in current density is normal and is attributed to higher electrode polarizations and ohmic loss at higher current density. During accelerated testing with extremely high HCL concentrations (5-40 ppm), some performance degradation was observed after 250-400 hours of operation. This was due to an increase in ohmic and anodic resistances, as confirmed by impedance spectroscopy.

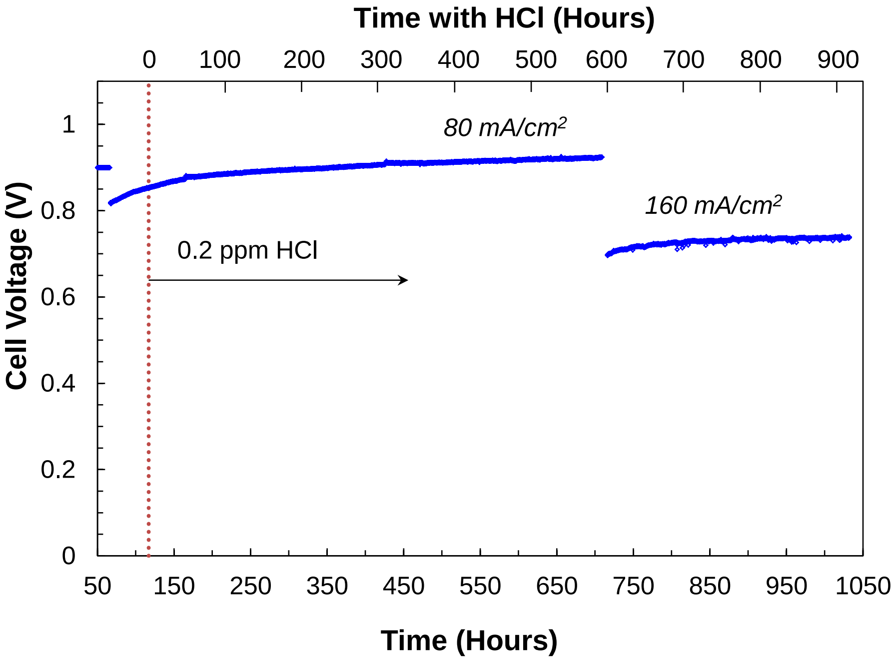


Figure 3-14. ECM Cell Performance (at 650°C) with Simulated Flue Gas Containing 0.2 ppm HCl

During post-test evaluation; {M}Cl crystals were found in the matrix, anode and anode current collector of the cells tested at high HCl concentration levels. This can explain a decrease of the ionic conductivity and higher ohmic losses observed at higher HCl concentration levels tested. This finding confirmed the modeling results, which indicated there is strong driving force for chloride from vapor to displace carbonate in the molten salt. It is believed that the {M}Cl formation and redistribution affected the anode and the electrolyte. As predicted by the modeling results, at expected HCl levels in flue gas, it would take >15 years of operation before significant electrolyte loss is observed.

Mercury: A button cell test was conducted to evaluate ECM cell performance with simulated flue gas containing 250 ppb Hg. Figure 3-15 shows the results of the 1100-hour test. No performance losses due to the presence of Hg^0 were observed over the whole test period. This is in agreement with the modeling data predicting no Hg^0 interactions with either of the cell components under the operating conditions. No changes in the ohmic or electrodic resistances were detected.

Since the presence of HCl is likely to lead to the $HgCl_2$ formation especially at lower temperatures, a test with simultaneous dosage of both Hg vapor and HCl was carried out to evaluate the effect of $HgCl_2$ on the cell performance. As shown in Figure 3-16, no effect of $HgCl_2$ (or 250 ppb Hg^0 and 0.2 ppm HCl) was observed during the 750-hour test.

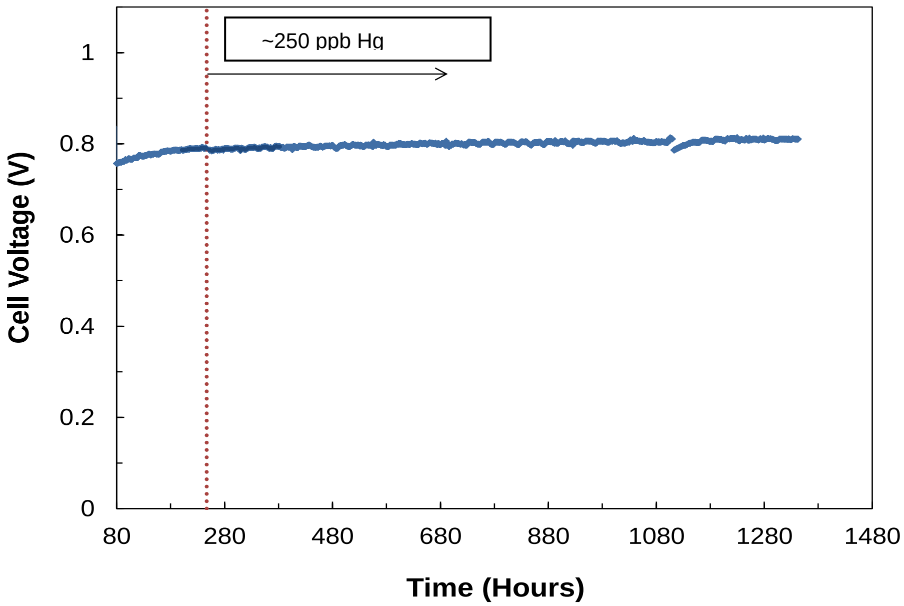


Figure 3-15. ECM Cell Performance at Constant Current Density of 110 mA/cm^2 (650°C) with Simulated Flue Gas Containing 250 ppb Hg: No degradation was observed during 1100- hour exposure

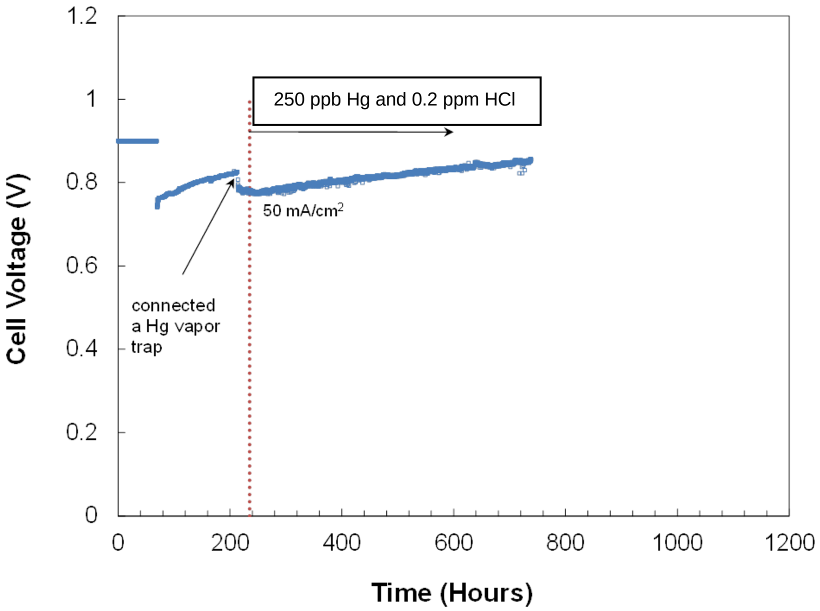


Figure 3-16. ECM Cell Performance (at 650°C) with Simulated Flue Gas Containing 250 ppb Hg and 0.2 ppm HCl: No degradation was observed during 750-hour total exposure

Selenium: An ECM cell was tested with simulated flue gas containing 10 ppb of selenium (SeO_2). The test was conducted at 160 mA/cm^2 and 650°C. Figure 3-17 presents the results. During the 860-h exposure to SeO_2, cell performance was very stable.

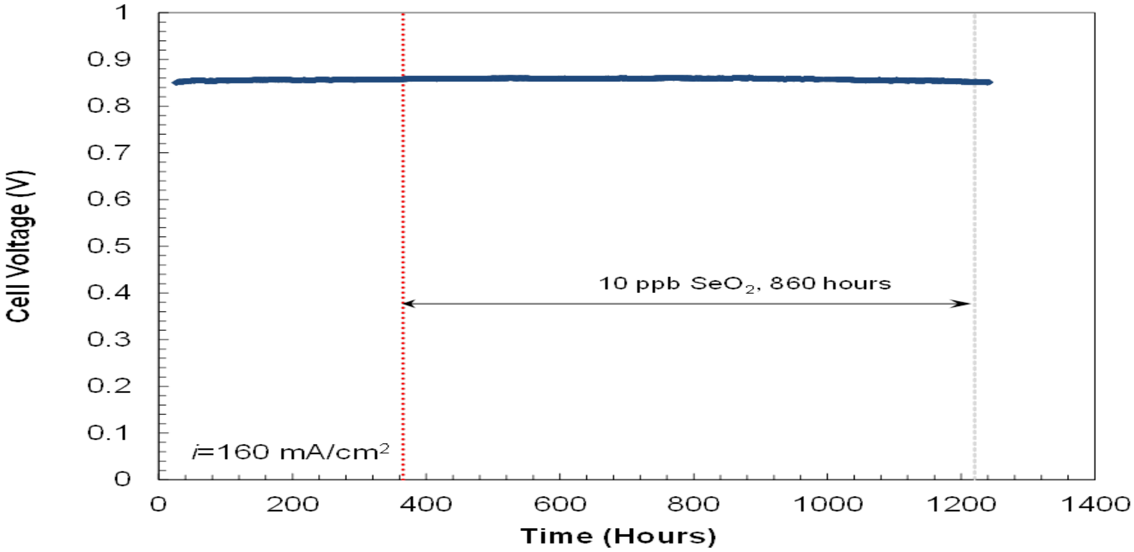


Figure 3-17. ECM Cell Performance at 160 mA/cm^2 (650°C) During Cathode Exposure to Simulated Flue Gas Containing 10 ppb SeO_2

3.3 Comparison with Contaminant Levels from Flue Gas Cleanup

Based on the experimental results, contaminant tolerance levels for the ECM were identified. The contaminant levels expected from the flue gas clean-up (polishing) subsystem were estimated and compared with the ECM tolerance levels. Table 3-1 provides a summary of the comparison. The contaminant evaluation and comparison with CEPACS plant flue gas polishing system output showed that the ECM tolerance levels are well above the contaminant levels expected in the ECM cathode feed gas (treated flue gas).

Table 3-1. Expected Flue gas Contaminant Levels Compared with Test Experience

Flue Gas Contaminant	Concentration in Cathode Inlet Gas After Polishing FGD, Estimated by AECOM		Highest Concentration Tested by PNNL, with low/no power degradation		Notes
SO_2	**0.18**	**ppmv**	**1**	**ppmv**	Performance losses due to short-term SO_2 exposure up to 40ppm were fully reversible
Se	**0.30**	**ppbv**	**10**	**ppbv**	No apparent degradation over 860 hours.
Hg	**0.08**	**ppbv**	**250**	**ppbv**	Expected form is predominantly elemental Hg. No apparent degradation over 1100 hours.
HCl	**12.7**	**ppbv**	**200**	**ppbv**	No apparent degradation over 900 hours.

4.0 SMALL AREA MEMBRANE TESTS

4.1 ECM (Membrane) Cell Performance Characterization

Single cell testing was performed using a 250 cm^2 cell in FCE's 7"x7" Cell Test Lab. After completing the qualification testing, system parametric testing was initiated. The testing was conducted using simulated clean flue gas. Cathode inlet gas composition from the CEPACS system developed was used as reference. The testing included three cell operating temperature conditions, four flow rate conditions, five reactant utilization conditions, and six cathode inlet O_2 concentration conditions. Cell performance (measured in volts), CO_2 flux (measured as current) and the percentage of CO_2 transferred (CO_2 utilization determined through mass balances based on the inlet and exit gas compositions) were characterized.

Testing began with setting the cell temperature to 620°C and characterizing the cell performance (voltage polarization as a function of current density) at various fuel and carbon dioxide (oxidant) utilizations. The anode and cathode flows were then increased (adjusted) to achieve the same utilizations and higher current densities. The cathode inlet O_2 and CO_2 concentrations were maintained at 8.3% and 9.9%, respectively. The utilizations and current densities used as the operating points are listed in Table 4-1. Five utilization conditions in the first column along with the current densities at each flow condition constitute the test matrix. The 100% flow condition corresponds to the oxidant and fuel flow rates required to achieve 92% CO_2 utilization and 68% fuel utilization, respectively at 110 mA/cm^2.

Table 4-1. Flow and Reactant Utilization Conditions Used for ECM Cell Performance Characterization

	Current Density			
	67% Flow	**90% Flow**	**112% Flow**	**135% Flow**
	mA/cm^2	mA/cm^2	mA/cm^2	mA/cm^2
30% Fuel & 40% CO_2 Utilizations	29.6	40.4	51.6	62.4
43% Fuel & 59% CO_2 Utilizations	44.8	60.8	77.2	93.2
52% Fuel & 71% CO_2 Utilizations	54.8	74.4	94.0	113.6
59% Fuel & 80% CO_2 Utilizations	62.4	84.4	106.8	128.8
68% Fuel & 92% CO_2 Utilizations	72.8	98.4	123.6	149.2

The percent carbon dioxide separated is equivalent to the CO_2 utilization in the ECM cathode. The carbon dioxide utilization was varied from 40% to 92%, the fuel utilization was varied from 30% to 68%, and the current density was varied from 29.6 mA/cm^2 to 149.2 mA/cm^2. Cell temperature of 620°C and cathode inlet oxygen concentration of 8.3% were used as the baseline conditions for testing. From these test data, constant flow polarization curves were generated for the cell. These cell performance characteristic curves are shown in Figure 4-1. Cell voltage decreases as the current density and utilizations increase.

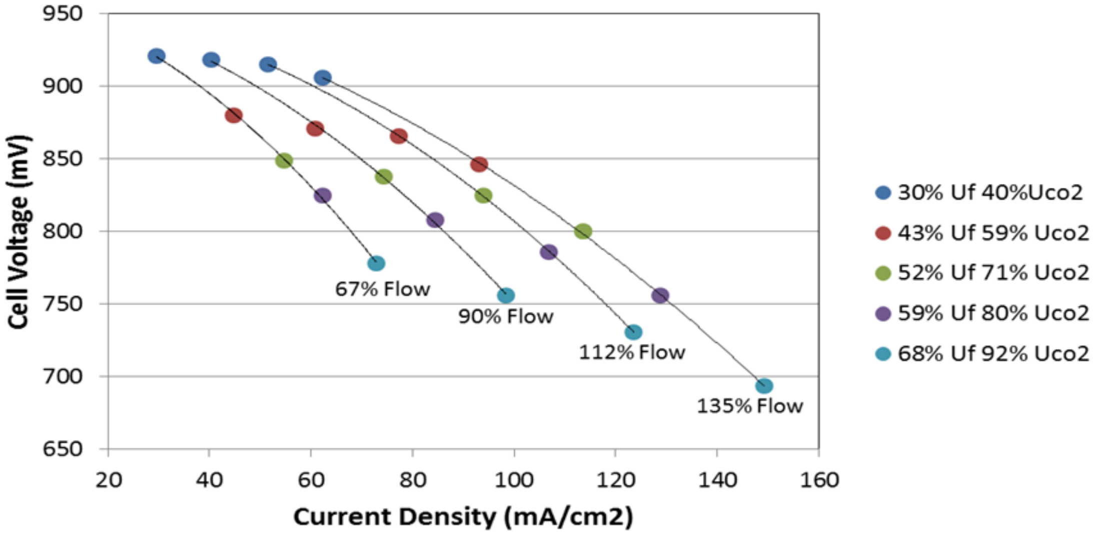


Figure 4-1. ECM Cell Performance Characteristics at 620°C Operating Temperature and 8.3% O_2 (9.9% CO_2) Cathode Inlet Gas Concentration

After the characterization at baseline conditions, cell performance at various cell temperatures and cathode inlet oxygen concentrations was examined. The cell temperature levels used in the study, in addition to the baseline condition of 620°C, were 600°C and 650°C. The temperature effect tests were conducted at the lowest and highest flow conditions of 67% flow and 135% flow. The cathode inlet oxygen concentration levels used in the study, in addition to the baseline condition of 8.3%, were 7%, 7.5%, 8%, 9%, and 9.7%. The last O_2 level condition (9.7%) was to simulate the gas composition of a recycle stream (in the CEPACS system). The oxygen concentration effects were studied at the three high flow conditions of 90% flow, 112% flow and 135% flow.

The test results associated with the temperature and oxygen concentration conditions are shown in Figures 4-2 and 4-3, respectively. The bar chart of Figure 4-2 shows the effect of cell operating temperature on cell performance at various reactant utilizations, for the highest flow condition of 135% flow. Based on Table 4-1, it can be seen that the comparison at 68% fuel utilization (Uf) and 92% CO_2 utilization (Uco_2) condition (right side of the bar chart) corresponds to the cell performance at 149.2 mA/cm^2. Whereas the comparison at 52% Uf and 71% Uco_2 condition (middle of the bar chart) corresponds to the cell performance at 113.6 mA/cm^2. Increase in cathode inlet O_2 concentration represents the extent of air supplementation of the flue gas. As the O_2 concentration is enhanced by the air addition, the CO_2 concentration decreases. The bar chart of Figure 4-3 shows the effect of O_2 Concentration on cell performance at various reactant utilizations, for 112% flow condition. The comparison at 68% Uf and 92% Uco_2 corresponds to the cell performance at 123.6 mA/cm^2. Whereas the comparison at 59% Uf and 80% Uco_2 condition corresponds to the cell performance at 106.8 mA/cm^2.

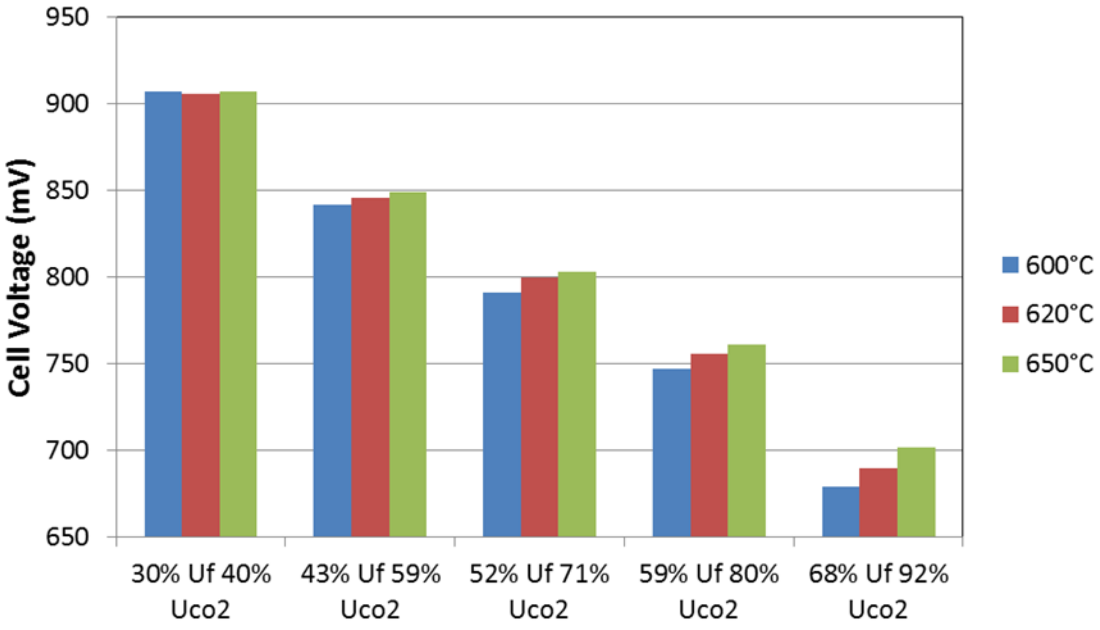


Figure 4-2. Effect of Operating Temperature on ECM Cell Performance with 8.3% O_2 (9.9% CO_2) Cathode Inlet Gas Concentration (135% fuel & oxidant flows)

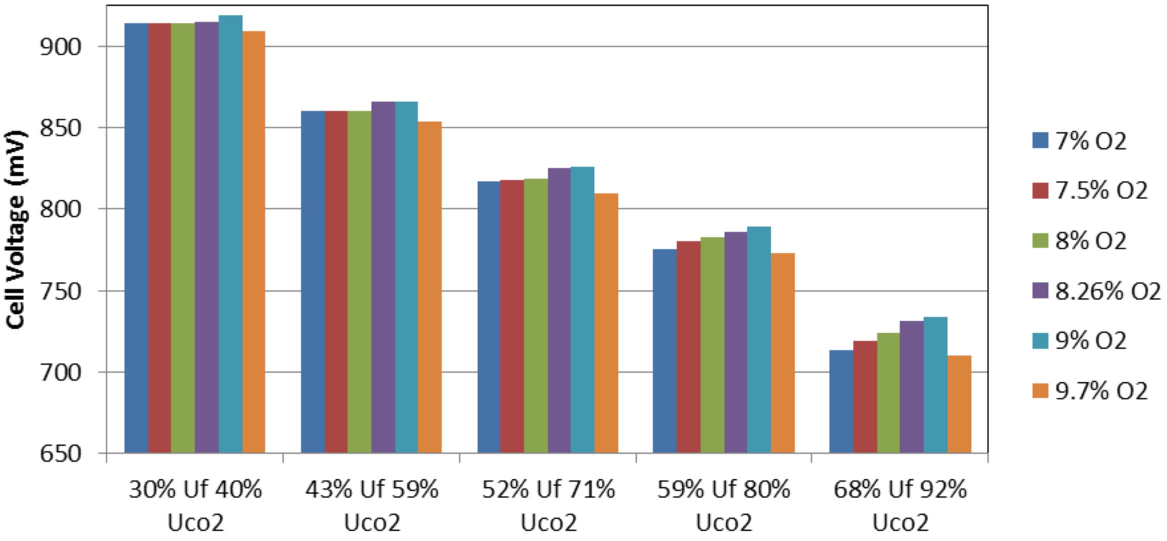


Figure 4-3. Effect of Cathode Inlet O_2 Concentration (air supplementation of flue gas) on ECM Cell Performance at 620°C Operating Temperature (112% fuel & oxidant flows)

From these results, it can be seen that the cell performance increases with the cell operating temperature. It is also apparent that as the cathode inlet oxygen concentration increases up to 9%, the cell performance increases. However, the performance decreases once the oxygen percentage reaches 9.7% (performance loss due to corresponding lower cathode inlet CO_2 concentration may be dominating). The resultant cell performance was affected by the operating temperature, cathode inlet oxygen concentration, and flow variations increasingly as the fuel and CO_2 utilizations were increased, as well.

From the parametric test data collected, performance at system design conditions (100% flow) was estimated by interpolation. From the interpolated performance data, the power production and CO_2 flux were calculated for baseline conditions as well as for optimized conditions (650°C, 9% O_2). The power produced at the baseline conditions as a function of current density is presented in Figure 4-4.

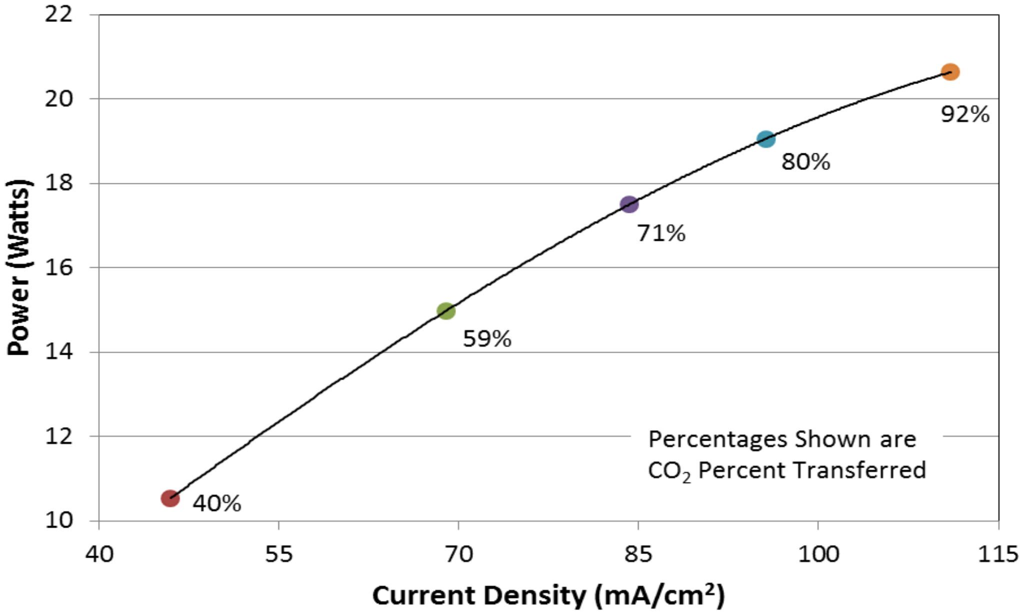


Figure 4-4. Power Generation and CO_2 Transfer (% of Cathode Gas CO_2 Separated) of ECM Cell as a Function of Current Density at System Design Condition of 100% flow

The chart demonstrates that as the current density increases, the power produced increases in a parabolic trend due to reduced cell voltages at higher utilizations and loads. As the current density increases, the CO_2 flux (and hence % of cathode gas CO_2 separated for a constant cathode feed flow rate) from the cathode to the anode of the ECM increases linearly. Figure 4-5 shows the relationships between these parameters and the improvement made by the optimized conditions.

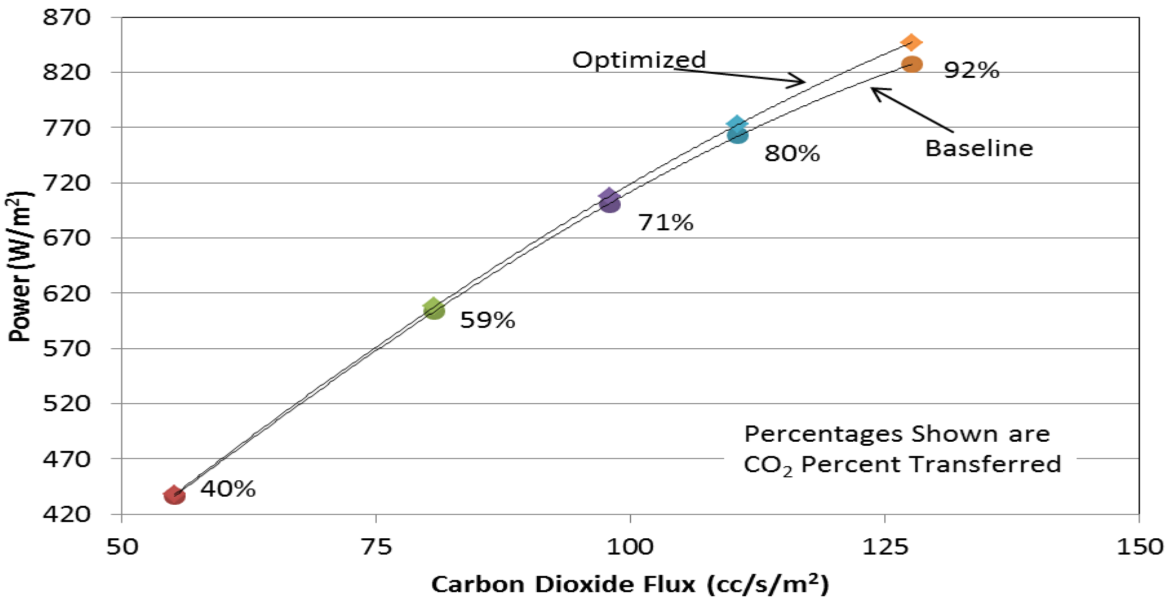


Figure 4-5. Power Generation, CO_2 Flux and % CO_2 Separation Relationships of ECM Cell and Improvement Made by Optimized Operating Conditions (650°C and 9% O_2)

As seen from Figure 4-5. , the power produced does increase with CO_2 flux. At the baseline standard conditions, the fuel cell is capable of producing 827 W/m^2 with a CO_2 flux of 127.7 $cc/s/m^2$. The optimized conditions resulted in a 2.4% increase in power production at the 92% CO_2 utilization condition.

A second cell (Cell 2409) was built and tested to check the performance reproducibility and conduct further parametric testing. In addition, this cell contained methane (CH_4) reforming catalyst on the anode side. This enabled testing with methane containing fuel. Initial testing included a performance comparison to previous cell (Cell 2317) at the standard system design conditions of 620°C cell temperature; fuel composition of 72.3% H_2, 18.1% CO_2 and 9.6% H_2O; and oxidant composition of 8.3% O_2, 10% CO_2, 12.6% water and balance N_2. The fuel and oxidant gas flow rates were set to achieve 68% fuel utilization and 93% CO_2 utilization (equivalent to percent CO_2 capture) at a current density of 110 mA/cm^2. Keeping the flow rates constant, the current was decreased in steps to generate a polarization curve covering a range of utilizations. Both cells performed comparably throughout the range of current densities and utilizations at the system design conditions. At the highest current density and utilization conditions, Cell 2409 had a voltage of 754 mV while Cell 2317 had a voltage of 752 mV.

Once it was determined that the cells had similar performance, some of the experiments previously performed were repeated in more detail and with more operational control through the use of mass flow controllers. This included examining the effect of O_2 concentration (variation representing the extent of air supplementation of the flue gas) on cell performance. Figure 4-6 presents the new test results. The test results correspond to 93% CO_2 utilization (U_{CO_2}, same as %CO_2 separation or capture level), 68% fuel utilization (Uf) and 110 mA/cm^2 conditions (equivalent to 128 $scc/s/m^2$ CO_2 flux). The system design condition has 8.3% (mole basis) oxygen concentration in the cathode inlet stream (increased from 2.3% O_2 in the flue gas by adding auxiliary air into the system). It can be seen that an oxygen concentration of 9% did not provide cell performance higher than that corresponding to the design condition of 8.3% O_2. The data showed that the cell performance begins to decrease as the oxygen concentration increases past 8.5%. This may be due to decreased CO_2 (the other oxidant needed for ECM operation) concentration resulting from increased air supplementation (CO_2 concentration decreases from 11.0 to 8.6% as O_2 concentration is increased from 7.0 to 10.0% in the test).

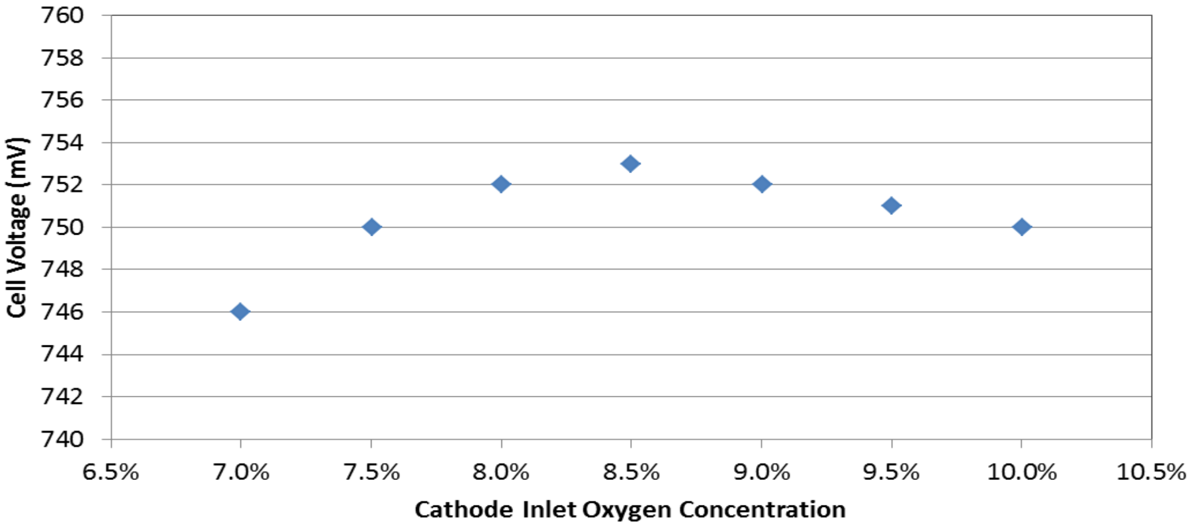


Figure 4-6. Effect of Cathode Inlet O_2 Concentration (air supplementation of flue gas) on ECM Cell Performance at 68% Uf, 93% Uco_2, and 110 mA/cm^2

As mentioned earlier, one of the primary differences between Cell #2409 and Cell #2317 was that Cell #2409 contained methane reforming catalyst on the anode side. Throughout previous testing (performed on Cell #2317), anode inlet composition of 72% H_2, 18% CO_2, and 10% H_2O vapor (80% H_2 – 20% CO_2 on dry basis) was used, which simulated fully-reformed methane fuel stream. To evaluate the cell performance with methane, a simulated partially-reformed methane fuel stream (representing preconverter effluent or Reforming Unit, RU, gas) was used. This stream consisted of 50% H_2, 25% H_2O, 12.5% CO_2, and 12.5% CH_4. As shown in Figure 4-7, the cell performance obtained with the two anode feed compositions was about the same. The test results correspond to 100% flow condition (at these flow rates 68% and 59% fuel utilization conditions corresponds to 110 and 95 mA/cm^2, respectively).

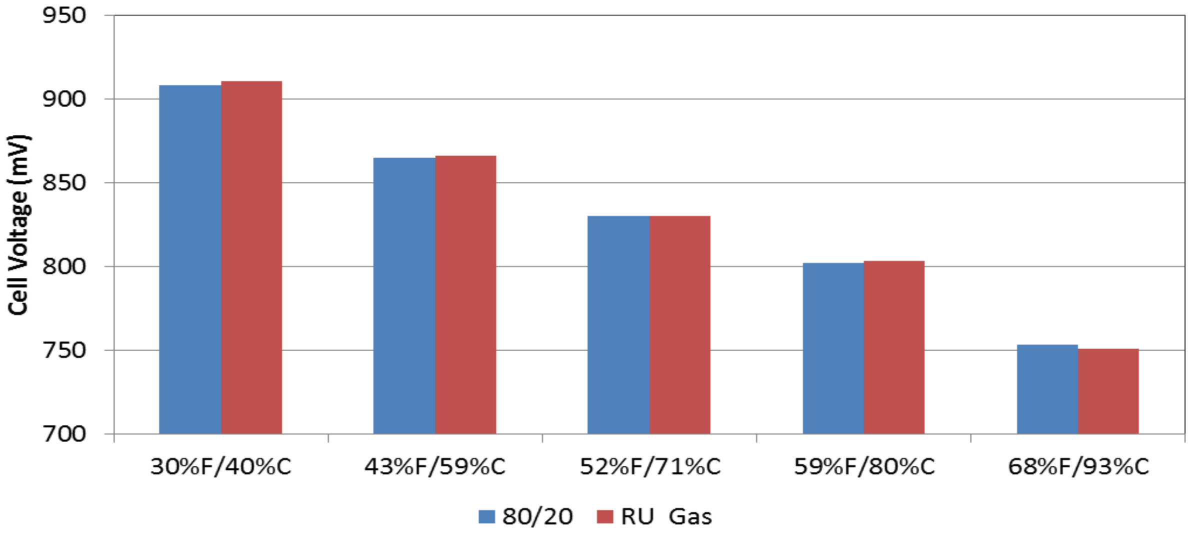


Figure 4-7. ECM Cell Performance Comparison Using Anode Feed Gas with Methane (RU gas) and without Methane (80/20)

Also, all prior testing was performed at a fuel utilization of 68%. System simulation results showed improved overall system efficiency at a fuel utilization of 75%. Therefore, an evaluation of the effect of fuel utilization (at a constant current density of 110 mA/cm^2 and a constant CO_2 utilization of 93%) on cell performance was performed. The results are shown in Figure 4-8. The change from 68% to 75% fuel utilization resulted in a 2% decrease in cell voltage.

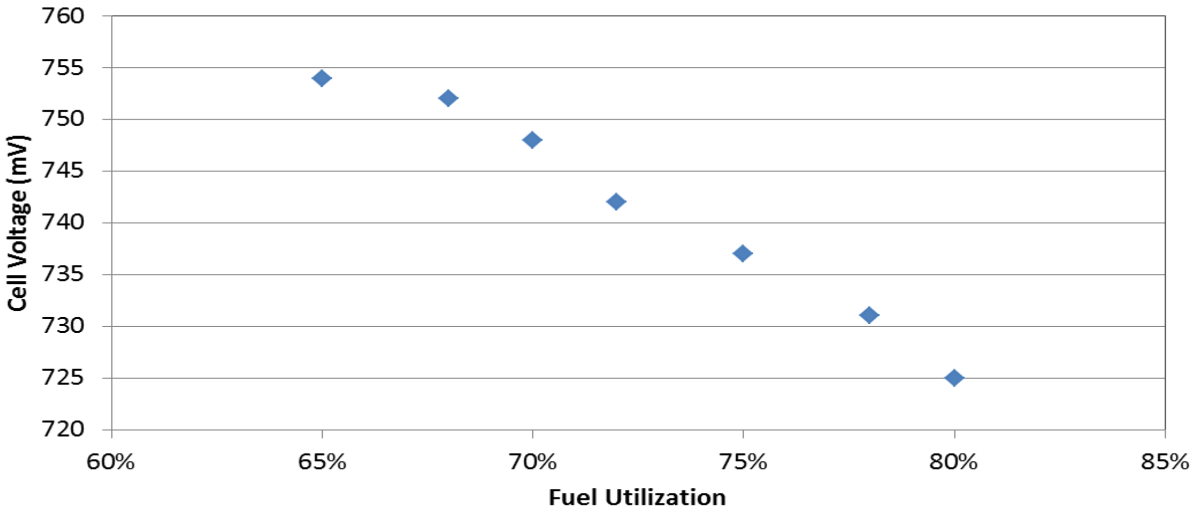


Figure 4-8. Effect of Fuel Utilization on ECM Cell Performance (at 110 mA/cm^2 and 93% CO_2 utilization)

The ECM cell performance was characterized at 75% fuel utilization (93% CO_2 utilization and 8.3% O_2 concentration in cathode feed) at a range of current densities. Figure 4-9 presents the constant utilization polarization characteristics.

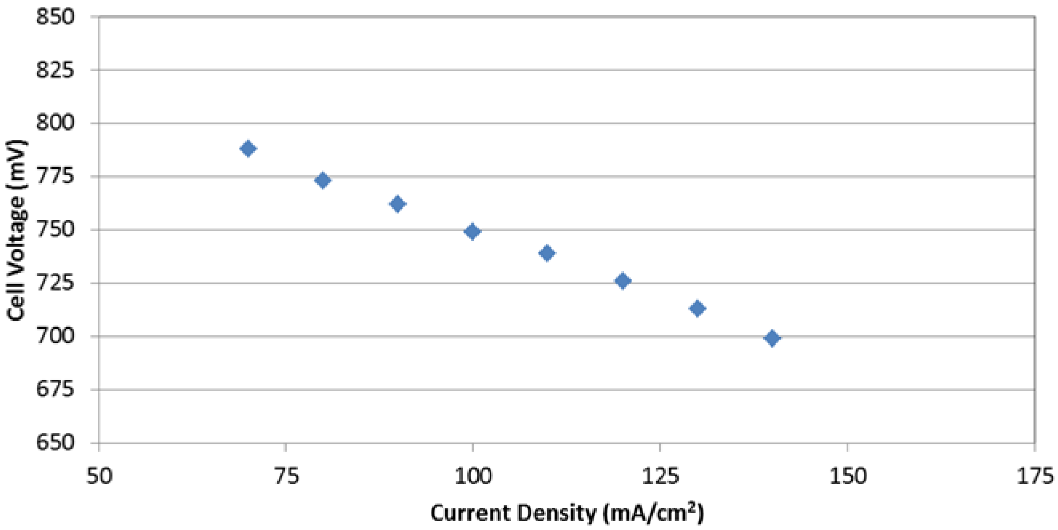


Figure 4-9. ECM Cell Constant Utilization Polarization Characteristics at 75% fuel utilization (93% CO_2 utilization)

Finally, to evaluate the long-term stability at the 75% fuel utilization, 93% CO_2 utilization and 110 mA/cm^2 condition with the system design condition gas compositions, steady state testing was conducted for over 2000 hours. Figure 4-10 shows that the cell performance remained stable with a mere 3.3 mV/1000 hours decay rate for the entire steady state hold period. The cell performance decay rate decreased to 1.4 mV/1000 hours for the last 1200 hours of the steady state hold.

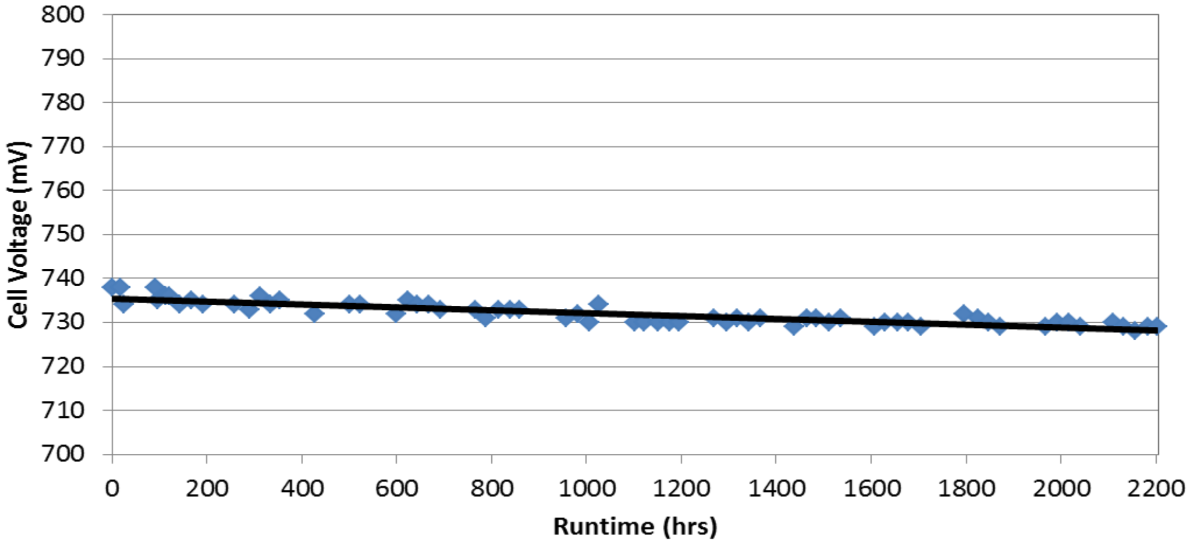


Figure 4-10. ECM Cell Performance during Steady State Testing at 75% Uf, 93% Uco$_2$ and 110 mA/cm^2

Despite the decrease in cell potential, the carbon dioxide flux through the membrane remained constant at 128 cc/s/m^2 throughout the 2200 hours of testing, as shown in Figure 4-11. At a constant current density, the CO$_2$ utilization and flux through the membrane remain constant.

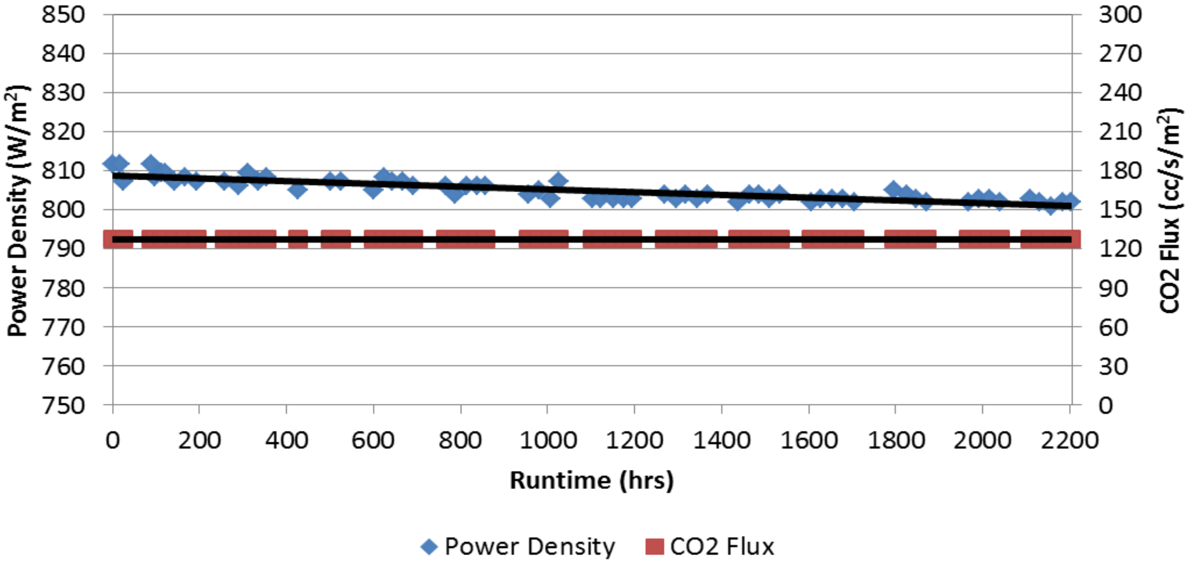


Figure 4-11. Power Density and Carbon Dioxide Flux during Steady State Testing of ECM Cell at 75% Uf, 93% Uco$_2$ and 110 mA/cm^2

4.2 Nitrogen Oxide (NOx) Related Testing

Testing was performed at FCE using a larger 250 cm^2 ECM cell to characterize the ECM's ability to reduce NO$_x$ emissions present in the PC plant flue gas. A compressed gas cylinder containing a gas mix with the composition of 500 ppm NO and balance nitrogen was used to introduce the contaminant into the clean cathode-in stream using a mass flow controller. The clean nitrogen (present in the clean cathode feed) flow was then decreased to maintain cathode inlet flow into the system constant. The resulting cathode inlet stream was also rerouted to bypass the cathode humidifier, to prevent the NOx from dissolving in the humidifier water (while bubbling through). NO$_x$ (as NO) concentration in cathode feed stream (flue gas) was increased from 50 to 216 ppm in five increments. For each concentration level, the cell operating current density was varied (from 48 to 110 mA/cm^2). NO$_x$ concentration in the cathode exit stream was measured at each operating point using a Thermo Scientific High Level Chemiluminescent NOx analyzer. Figure 4-12 presents the test results at 110 mA/cm^2. For 216 ppm cathode inlet NO concentration case, the cathode outlet NO$_x$ concentration decreased to 75 ppm. At least 70% of the NO$_x$ was removed at 110 mA/cm^2 operating condition for the range of cathode inlet NO concentration levels tested. Testing also included analysis of the anode exit stream, which confirmed that NO$_x$ was not transferred to the anode side of ECM cell.

It was further observed that NO$_x$ had no effect on the cell performance at the concentrations tested. It can be seen in Figure 4-13 that at each current density, the cell voltage varied very slightly between the six runs (concentration levels tested). This significant finding may eventually result in further reduction in the incremental cost of electricity associated with CEPACS system, if credits are given to elimination of the SCR in the coal power plant.

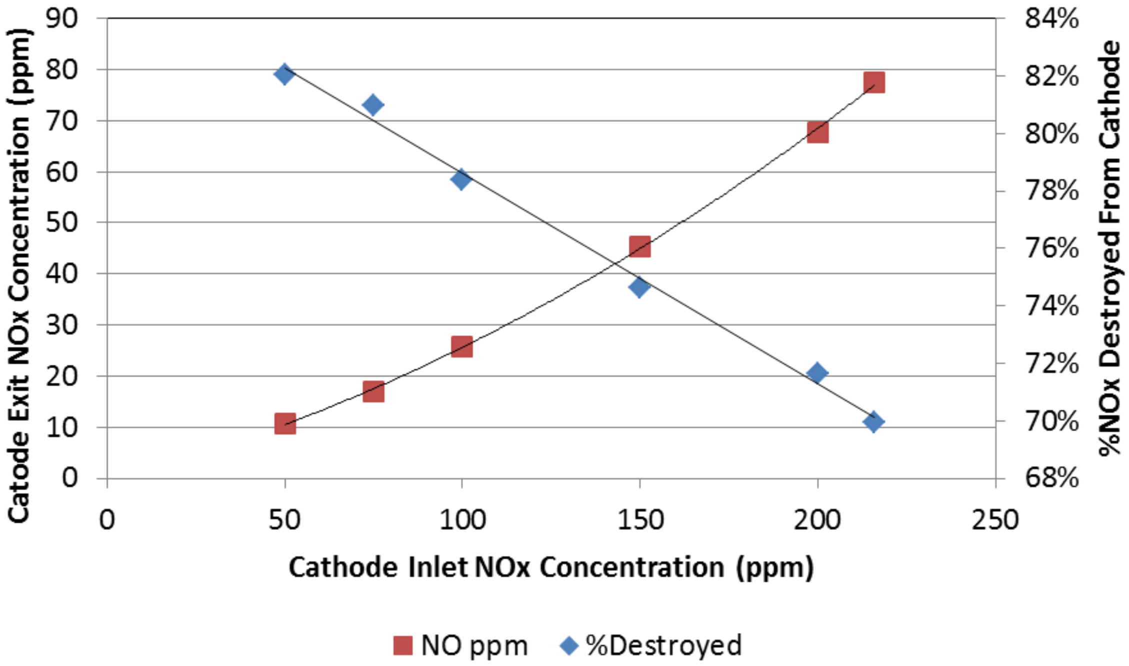


Figure 4-12. ECM Cell Cathode-Out Stream NO$_x$ Concentration and % NO$_x$ Destruction as Cathode-In NO$_x$ Concentration Increases at System Design Conditions

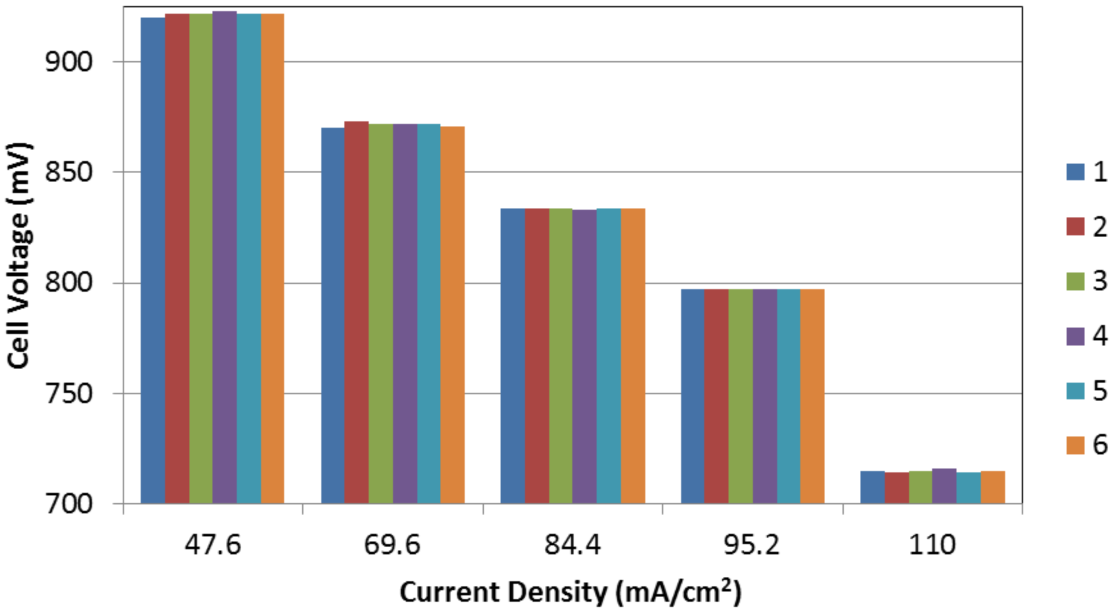


Figure 4-13. ECM Cell Performance during NO_x Tests (Operation at six NO_x levels)

5.0 BOP EQUIPMENT TECHNOLOGY GAP ANALYSIS

Technology Gap Analysis was conducted for the balance-of-plant (BOP) components (equipment) of the Electrochemical Membrane (ECM)-based CO_2 capture system. BOP equipment and related technologies were evaluated to check for commercial availability or need for a custom design. Technical and cost quotations for key equipment were solicited. Technology development, if required, and the extent of development to fill the technology gap were identified as an outcome of the evaluation process. Major components included Flue Gas Clean-up subsystem, chillers, CO_2 compressors and ECM stack interconnections (and enclosures). The findings are summarized in the subsections that follow.

5.1 Equipment Required for Flue Gas Clean-up

Traditional wet scrubber flue gas desulfurization systems are available from many large globally-known OEMs (original equipment manufacturers) such as Alstom, Advatech (Mitsubishi Heavy Industries), Steuler-KCH and Siemens. While wet-scrubber FGD systems are predominant, other technologies which are based on dry sorbents exist. Vendors with commercial offerings include: Hamon, Ducon, Macrotek, and Nol-Tec.

Both technologies have been developed for clean-up of the untreated flue gas from coal-fired power plants. These systems typically need to achieve concentrations in the tens of ppm for SO_2 and single digits of ppm for HCl. However, they have the potential, based on mass-transfer and reaction kinetics / equilibrium considerations, to achieve ppb levels of SO_2 and HCl after the clean-up.

AECOM has designed a proprietary two-stage system based on wet scrubber FGD technology, which is capable of achieving ppb levels of SO_2 and HCl. Advatech LLC (joint venture of URS and Mitsubishi Heavy Industries) has also developed a proprietary wet-scrubber FGD technology (to protect their proprietary amine system for carbon capture) which has been demonstrated at the 25 MW scale. It can process a treated flue gas (containing up to 10 ppm SO_2) to achieve ppb levels of SO_2.

FCE has investigated the potential of a single-step, dry sorbent process developed by Hamon to reduce SO_2 and HCl to the ppb level, as an alternative to the polishing, wet-FGD system designed by AECOM. This technology has been demonstrated on commercial-scale power plants as large as 625 MW, achieving single-digit ppm SO_2 levels from un-treated flue gas. FCE had discussions with vendors regarding the feasibility and sizing of a commercial-scale dry sorbent process. Preliminary estimates of the plant cost for the single-step, dry sorbent process (capable of ppb levels) developed by Hamon are $250 million, 120 percent higher than the estimate by AECOM for the two-stage, wet-scrubber FGD system.

In summary, flue gas clean-up technologies are commercially available and require moderate development for this application. Additionally, detectors to monitor concentrations of both SO_2 and HCl in flue gas are commercially available for the required range of concentrations.

5.2 CO_2 Capture and Compression Related Equipment

In the CEPACS plant, ECM-separated CO_2-rich stream is cooled, compressed and chilled to liquefy CO_2 which can be easily pressurized for sequestration or beneficial use. Information and findings for chillers and compressors is presented here.

Chillers: Key equipment in the CEPACS system for separating CO_2 from other gases present in the CO_2 – rich stream leaving the ECM anodes is the chillers. The chillers serve to liquefy the CO_2 at moderate pressures, allowing for easy down-stream separation of remaining gaseous hydrogen in the stream. The high-purity liquid CO_2 product is then pumped to supercritical pressure with minimal power requirement. A survey of vendors indicated that chillers are available from a large number of suppliers for applications requiring chiller capacity of up to about 5,000 RT (refrigeration tons). Table 5-1 includes a list of chiller suppliers.

Table 5-1. Chiller Suppliers

COMPANY NAME	ADDRESS	SIZE RANGE (Tons)
Energy Concepts Co LLC	727 Ridgely Ave Annapolis MD 21401	10-5000
GEA Refrigeration Technologies	3475 Board Rd York, PA 17406	100-1000
Petrogas Environmental Systems	PO Box 906 Angleton, TX 77516	100-1500
Thermax Inc.	21800 Haggerty RD Suite 112 Northville MI 48167	30-4000
Broad USA	401 Hackensack Ave Suite 503,v Hackensack, NJ 07601	30 -3000
York International Corp	631 South Richland Ave York, PA 17403	100-1400
Carrier	PO Box 4808 Syracuse, NY 13221	100-1700
The Trane Company	3600 Pammel Creek Rd La Crosse, WI 54601	100-1600

Larger capacity chillers (up to 18,000 RT) are needed for the CEPACS system. Two approaches were pursued to estimate the cost of the chillers in the CEPACS plant required for 550 MW PC plant. One approach was AECOM estimating chiller cost using their cost estimation data base. In the chiller costing conducted by AECOM, a 4 train (4x 25% capacity each) chiller system was assumed. The other approach was a funded study including chiller preliminary design and cost estimation, which was contracted to Energy Concepts.

Energy Concepts proposed a single train (e.g. centralized) chiller system. The design included two stages of chillers at two temperature levels. The first stage chiller design (moderate temperature) resulted in a single 4,400 RT unit, well-aligned with the size range of the vendor's commercial products (up to 5,000 RT). The Coefficient of Performance (COP) for the first stage chiller was 0.65, exceeding the process specification requirement supplied by FCE. The second stage (low temperature) requiring 18,000 RT capacity would be too large as a single unit. Therefore, the vendor quoted a modular design of 5 units, 3,600 RT each, based on existing commercial offerings. The Coefficient of Performance (COP) for the second stage chiller was 0.45, meeting the process specification supplied by FCE.

In summary, the chillers required by the CEPACS system are based on commercially available technologies and performance levels. As the refrigeration tonnage requirement is large, a modular approach to utilize multiple parallel trains of commercially-proven size equipment was taken. The estimated chiller costs represented only ~3% of the Total Plant Cost, indicating economic viability for the process.

Compressors: To verify the technical and economic feasibility of the compressor system, vendor technical/cost proposals were solicited from five global compressor OEMs, including:

- Siemens
- GE Oil & Gas
- MAN Diesel & Turbo
- Elliott Group (Ebara Corporation)
- Dresser-Rand

AECOM received proposals from three. The compressors in these proposals comprised state-of-the-art machinery, with a broad range of applications, including: industrial gases, petrochemicals, hydrocarbons, power generation, furnace gas recovery (FGR), ammonia, etc.

The performance of the proposed compressor systems was well-aligned with the 85% efficiency assumed in the process model developed by FuelCell Energy. The efficiencies ranged from 78.5 to 89.5 percent. Hydrogen embrittlement is generally a concern when streams contain hydrogen at elevated partial pressures. When soliciting the technical proposals, all of the OEMs were informed of the hydrogen content in the CO_2-rich stream. The compressors in all of the proposals are suitable for operation in this environment.

Of the three cost proposals received, cost estimates from MAN and Elliott were within 1.7 percent of each other, providing a high degree of confidence in the accuracy of the cost. The proposal from Siemens was substantially higher, as the needed capacity far exceeded their largest compressors and required multiple, parallel trains. The CO_2 compressors are a small fraction (~4%) of the overall CEPACS cost, while they account for an estimated ~69% of the CEPACS system auxiliary power load (~11% of the gross power). Future incremental improvements in both efficiency and capital equipment costs will have tangible impacts on the overall economics of the CEPACS system. Emerging technologies (e.g., Ramgen) may offer benefits in one, or both, of these attributes.

In summary, CO_2 compressors are commercially available and require no development for this application.

5.3 ECM Stack Interconnections

Ducting and piping required to interconnect the ECM stacks and route gases throughout the CEPACS plant represent a significant portion of the system capital cost (up to 10%). To reduce the cost of ECM interconnections (as estimated in BP1), the following approach was taken:

1. Solicit vendor quotations and design guidance
2. Develop a larger ECM stack module/enclosure building block and associated plant layout to reduce the amount and length of interconnections

The technical availability of very large custom-fabricated ducting has been demonstrated in power industry and petroleum-processing applications. However, custom field-erected ducting is expensive. Through vendor contacts and quotations, it was learned that factory-fabricated ducting, readily available in diameters up to 40”, is much cheaper than field-erected ducting. Therefore, multiple smaller ducts result in reduced cost compared to a single large custom duct. Also fittings may account for 50-80% of ducting material costs. Therefore, straight runs of ducting were selected to reduce cost. The plant layout (presented later in this section) was developed utilizing the above-mentioned design guidance. The layout designs presented in this section were developed during BP2 of the project as an interim update on the BP1 plant design presented in Section 1.0. Section 8.0 describes the updated plant layout developed during BP3 of the project.

To reduce the amount and length of interconnections, a larger ECM stack enclosure building block was developed. The larger building block enclosure housed 200 stacks in a single enclosure compared to four (4) stacks per enclosure in the BP1 design. In addition to reducing the interconnection costs, this design strategy also leveraged economies of scale resulting in lower normalized enclosure costs. A description of the CEPACS plant design, including stack enclosures and interconnections, suitable for 90% CO_2 capture from a reference 550 MW PC plant follows.

To reduce plant capital costs, and decrease heat losses and pressure drops, the high (operating) temperature BOP equipment was located proximate to the 10 ECM enclosures as shown in Figure 5-1. This approach greatly reduced the lengths of high temperature piping and ducting. Conversely, the low (operating) temperature BOP equipment was centralized in a single area. The two areas were connected by a small number of headers providing the reactants and the required utilities and removing the exhaust gases.

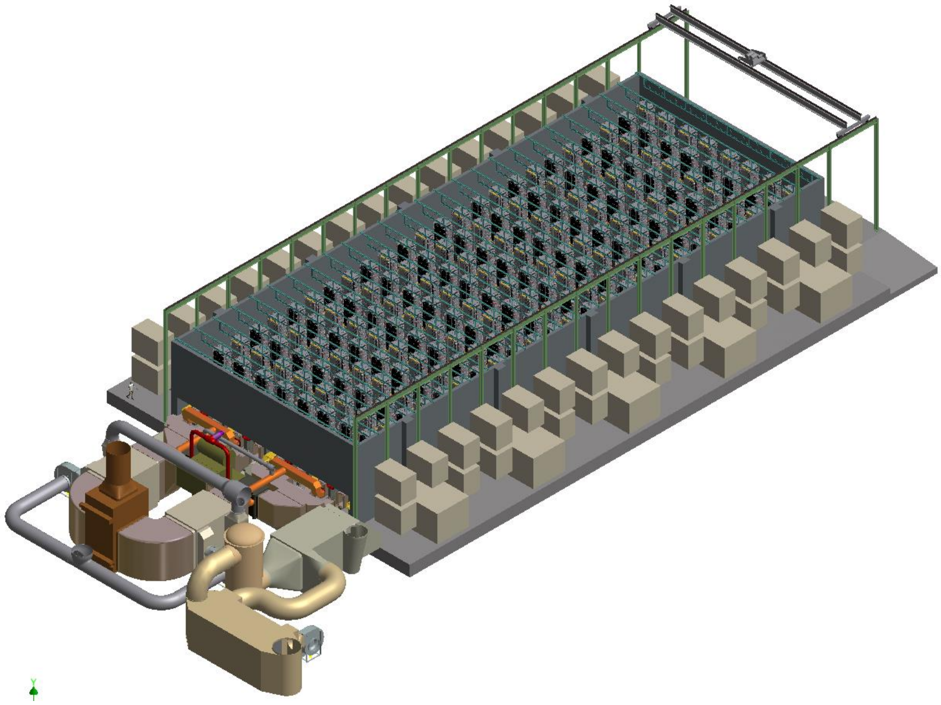


Figure 5-1. **General Arrangement of an ECM enclosure and its “hot” BoP equipment**

The general arrangement of the ECM enclosures and associated “hot” BOP equipment is shown in Figure 5-2, along with the piping for the distribution of coal plant flue gas to and collection of the CO_2-rich anode exhaust gas from the plant sections (ECM enclosures). The remainder of the balance-of-plant equipment such as the compressors and chillers was centralized and located in an area to the side of the ECM enclosures and opposite to the FGD unit.

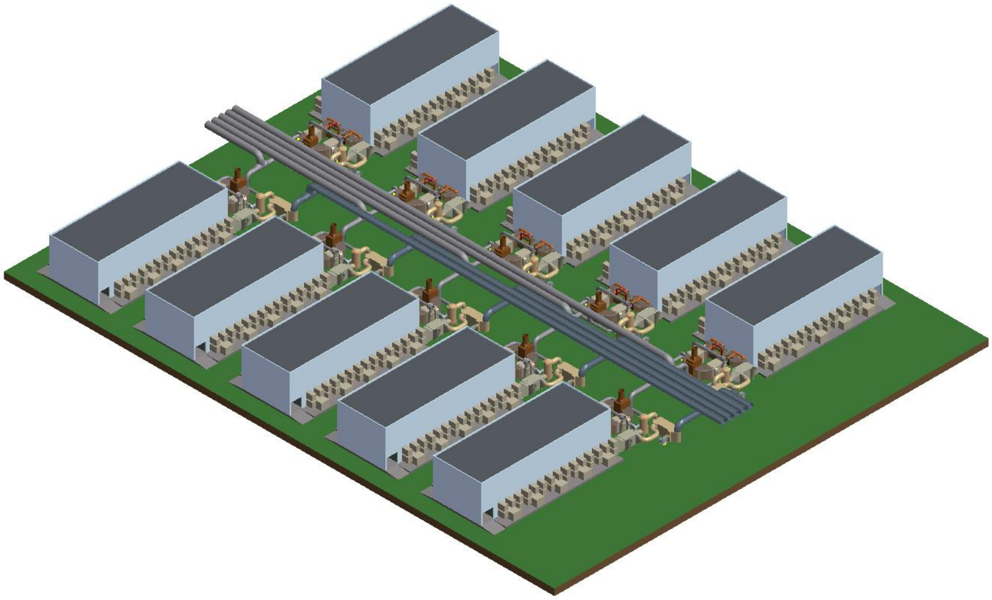


Figure 5-2. **General Arrangement of ECM Enclosures, BoP Equipment and Piping**

The low temperature piping and ducting (including the headers) was designed so that standard, commercially-off-the-shelf (COTS) products can be used. Carbon steel was used throughout, except where carbonic acids levels required the use of stainless steels. A potential, low-cost product is Spiral Buttweld Pipe manufactured by Naylor in diameters up to 102 inches (8.5 feet) using standard gauge sheet metal. Most of these pipes and ducts were not insulated, further reducing their cost.

The high temperature piping and ducting were designed in conjunction with the BOP equipment to reduce plant capital and operating costs. This required the high temperature piping and ducting to be custom fabricated, however, it facilitated the close placement of the high temperate BOP equipment, as illustrated in Figure 5-2. This, in turn, minimized the length of the ducting, reducing heat losses and pressure drops. It also facilitated the splitting of some heat exchangers into two smaller heat exchangers which are easier to transport and install.

The high temperature piping and ducting will be fabricated from carbon steel wherever possible; otherwise, stainless will be used. Two vendors familiar with fabricating similar high temperature ducting for the Power Generation and Chemical Process Industries are Nadine Corporation and PSP Industries.

The piping and ductwork cost estimate was developed utilizing the pipe sizes estimated. Several assumptions/good design practices were employed to determine the cross sectional area of the ductwork and estimate the material costs. Velocities in the ductwork were assumed to be limited to 60 ft/sec (18.3 m/s). This velocity is typical for FGD ductwork systems, as well as compressed air piping, to stay below acoustic velocity limits and avoid excessive noise generation in the ductwork and/or pipes.

Overall, the piping and ducting for the CEPACS system uses a combination of readily-available, commercial-off-the-shelf (COTS) and custom-fabricated products. Significant conceptual design optimization has resulted in reduced ECM interconnection and enclosure costs.

6.0 TECHNICAL ENVIRONMENTAL HEALTH AND SAFETY REVIEW

An Environmental Health and Safety (EH&S) review of the ECM-based CO_2 capture and compression process was carried out in Budget Period 3 of this project. The review included the identification and quantification of the emissions from the CEPACS system. Properties of the waste including reactivity, flammability, corrosivity, and toxicological effects were discussed along with methods for minimization/elimination where applicable. The study was conducted for the ECM-based carbon capture system of size suitable for a 550 MW PC power plant. The PC plant design specified in “Case 9” of the DOE - NETL report “*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2” wa*s used as a reference. The CEPACS system (carbon capture system) employs FCE’s Direct Fuel Cell technology which is commercialized for stationary power generation applications. The CEPACS plant utilizing ECM technology captures and compresses >90% of CO_2 from the flue gas of the reference 550 MW (net AC) PC Rankine Cycle Plant.

The CEPACS system emissions include gaseous emissions during normal operation, such as CO_2, SO_X, NO_X, particulate matter and Hg. The major liquid discharge is boiler and cooling water blowdown. The chemicals used in the treatment of boiler water and cooling water were considered along with EH&S related information. Ancillary emissions from the CEPACS system include periodic discharges and solid wastes such as sorbents and catalysts. Incidental emissions include refrigerants used in chillers and lime used in the polishing FGD. The chillers in the CEPACS system use ammonia as the refrigerant. The related results and information are summarized below.

6.1 Gaseous or Air Emissions

The gaseous or air emissions were estimated for the PC plant (Case 9 of DOE-NETL Report) equipped with the CEPACS system based on a total net power output of 971 MW and are presented in Table 6-1. The reported emissions include the greenhouse gas (CO_2) emissions and emissions of the individual pollutants such as SO_2, NO_x, Hg and Particulates.

Table 6-1. Air Emissions for 550 MW PC Plant Equipped with CEPACS System

	kg/GJ (lb/MMBtu)		Tonne/year (ton/year) 85% CF		kg/MWh (lb/MWh)	
CO_2	5.2	(12.2)	349,442	(385,190)	43	(95)
CO_2[1]					48	(107)
SO_2	0.0002	(0.0006)	16	(18)	0.002	(0.004)
NO_x	0.006	(0.014)	414	(456)	0.051	(0.113)
Particulates	0.002	(0.004)	113	(124)	0.014	(0.031)
Hg	0.0000	(0.00)	0.020	(0.022)	2.43E-06	(5.37E-06)

[1] CO_2 Emissions Based on Plant Net Output

Carbon dioxide emitted with the plant exhaust (ECM cathode side) represents CO_2 in the flue gas which is not captured by the ECM stacks. This is reported as a mass flow per net power. The percentage of CO_2 captured exceeds the 90% capture level minimum. Figure 6-1 shows the normalized emission rates of CO_2 calculated for the three (system) cases defined in Table 6-2. Case 3 (PC + CEPACS) has the lowest CO_2 emissions of all the cases. This is due to the design criteria to capture 90% of the carbon which enters the CEPACS system. Since the CEPACS system generates additional power, the CO_2 emissions (normalized by total plant output) are ~60% lower than for Case 2 (PC plant equipped with competing amine absorber based CO_2 capture technology). Case 1 relates to emissions of the baseline PC plant not equipped with the CO_2 capture system.

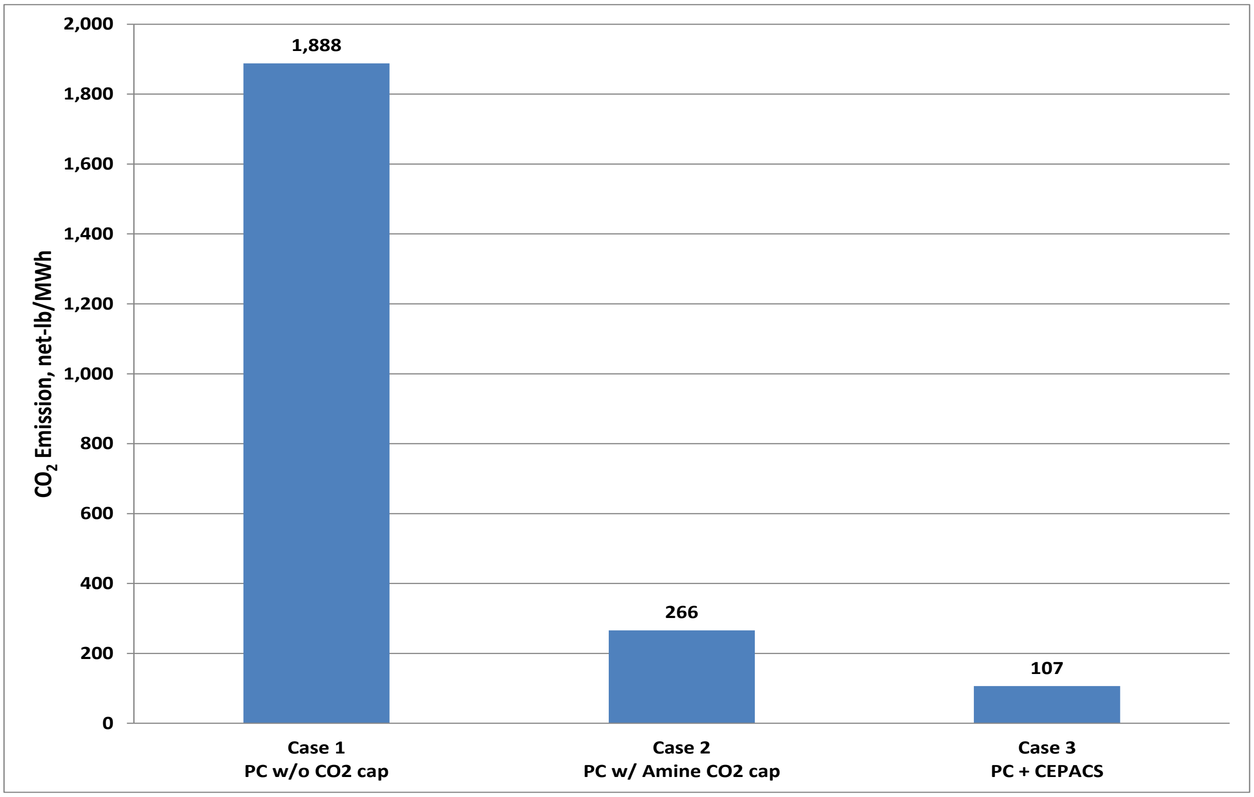


Figure 6-1. CO_2 Emission Rates

Table 6-2. Case Summary

Case ID	DOE/NETL-2010/1397 Case ID [2]	Unit Cycle	MW gross	MW net	Boiler Technology	Sulfur Removal/ Recovery	CO_2 Separation
1	9	PC	583	550	Subcritical PC	Wet Flue Gas Desulfurization (FGD)	--
2	10	PC	675	550	Subcritical PC	Wet FGD	Amine Absorber
3	--	PC + ECM	1086	971	Subcritical PC	Wet FGD	ECM-based CEPACS

The primary and the polishing SO_2 scrubber units (FGD) reduce the SO_2 level in the flue gas (fed to the ECM cathode) to 0.3 ppmv. The SO_2 which is not removed by the primary (1st) Wet Limestone Scrubber and the CEPACS Polishing Wet Limestone scrubber (2nd) is emitted to the environment. All sulfur is reported as SO_2. Sulfur in the natural gas fuel (supplemental fuel for ECM anode feed) is adsorbed and captured by the fuel desulfurizer sorbent.

The NOx emissions correspond to the flue gas NO_2 and NO content which is not removed/converted by the ECM stacks. ECM testing results have shown that during normal power operation >70% of the flue gas NO_x content is transferred across the membrane to the anode and reduced. No NOx emissions are created within the CEPACS system. The overall emission specific flow rate of 0.051 kg/MWh is a fraction of the emission limit of 0.339 kg/MWh set by the US EPA new source performance standards (NSPS) Mercury and Air Toxics Standards (MATS).

Particulate emissions account for the PC Plant flue gas particulate matter not removed by a combination of the PC plant baghouse and the two Wet Lime scrubbers (FGD) in series. These solids are removed at 99.8% removal efficiency. The remaining particulates exit with the CEPACS plant exhaust stream. The emission level is a fraction of the MATS limit of 0.041 kg/MWh.

Figure 6-2 shows the normalized emission rates of SO_2, NOx, and particulate matter (PM) calculated for the three (system) cases. NSPS MATS limits are also shown in the Figure. Case 3 has the lowest emission rates of SO_2, NOx, and PM of all the cases studied. Case 3 emission rates are significantly lower than those required by the NSPS MATS rules. Emissions of NOx and PM for Case 2 exceed the MATS requirements.

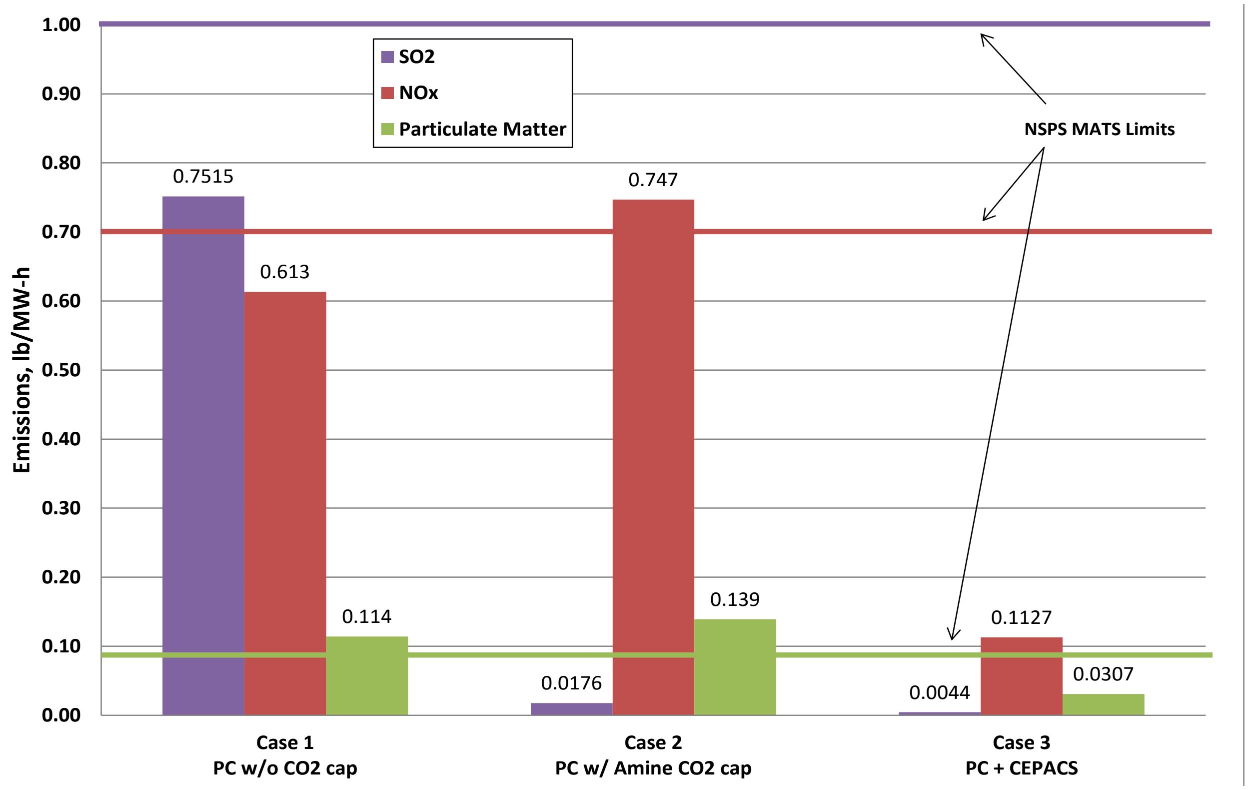


Figure 6-2. SO_2, NOx, and Particulate Emission Rates

Mercury (Hg) emissions represent the mercury content in the PC Plant flue gas which is not scrubbed out by the Primary wet lime scrubber (FGD) unit. This primary scrubber removes 90% of the Hg in the flue gas feed by wet lime impingement and conversion. No credit is taken for Hg removal in the downstream polishing scrubber (FGD). The mercury passes through the ECM cathode side, as verified by ECM contaminant evaluation testing. Figure 6-3 shows the normalized emission rates of mercury calculated for the three cases. Case 3 has lower Hg emissions than Cases 1 and 2, because a portion of the net power generation is fueled by natural gas which contains negligible amount of Hg.

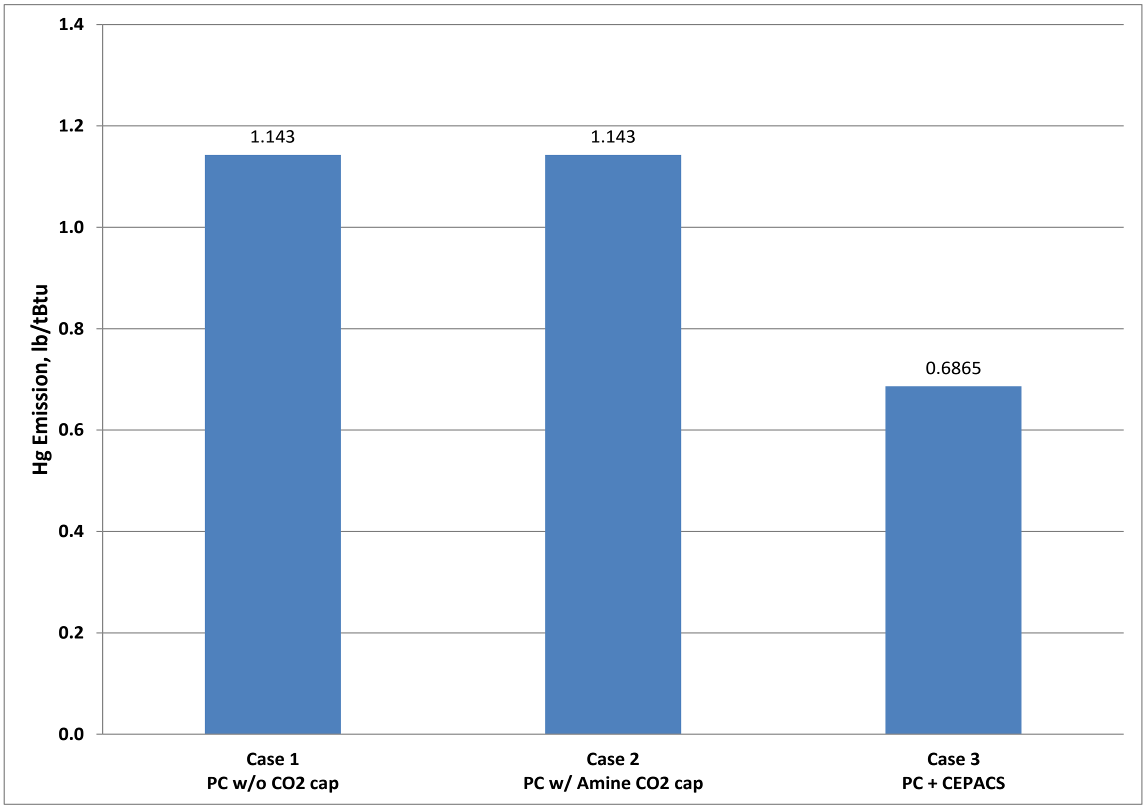


Figure 6-3. Mercury Emission Rates

6.2 Liquid (or Water) and Solid Emissions (including FGD Unit Related)

In this subsection, liquid or water emissions of the CEPACS plant in general are reviewed and discussed first, followed by the discussion of the liquid and solid emissions specific to the Polishing FGD unit.

Plant Water Balance and Liquid Emissions (Discharge)

Table 6-3 shows the overall water balance for the 550 MW PC plant equipped with the CEPACS system (Case 3) and includes the water usage of the base PC plant. The table also presents a break-down including the major items. The CEPACS system (when in operation) is a net producer of clean process water which is sent to the PC plant for process use. CEPACS plant internal water consumption includes boiler water (steam) required for ECM fuel feed and quench water used upstream of the shift converter (LTSC).Because the CEPACS system (including the Polishing FGD) will be processing the PC plant flue gas, the heat duty /size of the (PC plant) cooling tower will be increased. The cooling water is used in CEPACS heat exchangers and chillers.

The primary exit water stream from the Cooling Tower will include the additional blowdown water proportional to the above-mentioned CEPACS heat exchanger cooling water loop load. A large fraction of the cooling tower blowdown water (500 gpm) is utilized as feed water to the Polishing FGD scrubber. This feature serves to reduce the net water withdrawal of the CEPACS system and also reduce the plant liquid discharge. Most of the scrubber feed water is consumed in making slurry and gypsum. The remaining portion of the cooling tower blowdown water (which is a small fraction of the cooling tower feed water) goes to the Waste Water clarifier.

Table 6-3. Water Balance for 550 MW PC Plant Equipped with CEPACS Plant

	Raw Water Withdrawal, m^3/min (gpm)	Process Water Discharge, m^3/min (gpm)	Raw Water Consumption, m^3/min (gpm)
Primary & Polishing FGDs	3.9 (1,017) Note 1	1.9 (500)	3.9 (1,017)
BFW Makeup	0.3 (74)	0.0 (0)	0.3 (74)
CEPACS Process	-1.4 (-381) Note 2	0.0 (0)	-1.4 (-381)
Cooling Tower	30.8 (8,120)	5.2 (1,367)	23.7 (6,253)
Totals	**33.6 (8830)**	**7.1 (1,867)**	**26.5 (6,963)**

Note 1: Polishing FGD utilizes blowdown from the cooling tower as feed water; no raw-source water is used.

Note 2: Negative numbers indicate water exported from (or produced by) the system

Polishing FGD Unit and Related Liquid and Solid Emissions (Discharge and Solid Waste)

A discussion of the polishing FGD unit (which is part of the CEPACS system) related emissions and their handling is presented here along with the related process description. The polishing FGD will be located after the primary FGD scrubber. The polishing FGD will be a wet scrubber using limestone slurry to remove SO_2 from the flue gas. Gypsum is produced in a reaction tank which is a part of the FGD unit. A bleed stream of gypsum slurry must be removed from the reaction tank to control density. The stream is sent to a dewatering system (discussed separately below) where the solids are directed to the same location as the gypsum produced in the primary FGD system. The limestone feed system and the gypsum dewatering system for the Polishing FGD are common to the primary FGD equipment.

The FGD water discharge is estimated to be 500 gpm. There will not be any evaporation in the polishing FGD, since it will be located downstream of the primary FGD and the flue gas is, therefore, cooled and saturated before reaching the polishing FGD. The discharge is sent to wastewater treatment (WWT), which helps control chlorides and heavy metals concentrations in dewatered gypsum byproduct (discussed further later in the section).

Dewatering System and Handling of Recovered Water: Dewatering mentioned above can be performed in several different ways. The common technologies used are:

- Horizontal Belt Filters (often in conjunction with hydrocyclones),
- Belt or Drum Filters
- Centrifuges.

The chosen technology depends on the requirements for the gypsum (calcium sulfate) byproduct. If wallboard quality gypsum is to be generated (i.e., less than 10% moisture), hydrocyclones followed by horizontal belt filters are used. If the produced gypsum is to be landfilled (~15% moisture), drum filters are used.

Although the polishing scrubber is as large as the primary scrubber (in order to achieve the required mass transfer and removal percentages), the polishing scrubber is removing a much smaller mass of SO_2 (2% of the SO_2 is removed in the polishing scrubber relative to the primary scrubber). Accordingly, the polishing scrubber adds only a small amount of slurry to the existing dewatering system (of the PC plant).

Water recovered from the dewatering system is commonly settled and ultimately discharged through a plant's wastewater outfall. Some contemporary designs employ zero liquid discharge (ZLD) systems where all liquid from the dewatering system is returned to the FGD scrubber for use in limestone grinding operation and/or as scrubber make-up water. However, returning all the water to the FGD system will increase the concentration of chlorides in the water (chlorides present in the flue gas are water soluble), making it necessary to evaluate the materials of construction for the FGD system. Oftentimes for ZLD systems, the FGD equipment must be modified and higher grade metal alloys used for construction that can tolerate higher chloride concentrations. Chlorides can be controlled through the removal of water from the system and using waste water treatment facilities to treat and discharge a portion of the dewatering system effluent, removing chlorides along with the water.

Overall, water recovered from the FGD slurry is largely recycled back to the FGD system, with added design considerations for higher chloride concentrations. The water generated from the polishing scrubber is a small fraction (2%) of the water generated from the primary scrubber.

Heavy Metals and Their Disposal: The majority of heavy metals from the coal will be found in the fly ash, which is removed from the flue gas via particulate control (part of the PC plant). A percentage of the heavy metals will pass through particulate control to the primary FGD scrubber. Mercury is often found in scrubber slurries. Since oxidized mercury is water soluble, the majority of the oxidized mercury will be captured in the primary scrubber. Some unique characteristics may limit the removal of certain species of selenium. Hence, some selenium may pass through the primary scrubber to the polishing scrubber. In general, heavy metals are not anticipated at problematic concentrations in the polishing FGD scrubber because of their removal via particulate control, the water solubility of heavy metals, and the low relative removal level of the polishing scrubber compared to the primary FGD scrubber.

In the event heavy metals originating from the coal used in PC plant are found in the FGD gypsum product, the resultant gypsum cannot be sold for reuse (i.e., wallboard manufacture) but instead must be landfilled appropriately.

Chloride-containing Waste Water and Related Handling: Chlorides present in the flue gas are water soluble and will be removed almost completely by the FGD scrubbers. In general, FGD systems do not generate high chloride waste streams (> 10,000 ppm). However, if high chloride levels resulting from severely reduced waste water discharge flow from the FGD scrubbers (approaching or reaching ZLD) are problematic, then additional treatment process using a brine concentrator or waste water evaporator would need to be employed.

Limestone Preparation and Related Handling: The polishing FGD scrubber will use a small slipstream of limestone reagent from the primary FGD scrubber (~2%). Therefore, no additional hazards are introduced by the use of limestone in the polishing scrubber. For the primary scrubber, limestone would be purchased pre-crushed. A majority of FGD systems grind limestone rock on-site. The most common grinding technique is to use a ball mill to reduce the limestone from a ~¾" rock to a 30 wt% solid (in water) limestone slurry (where 90% of the solid material can pass through a 325 mesh (44 micron) screen). The slurry is pumped to the FGD system. In addition to ball mills; tanks, hydrocyclones and pumps are a part of the limestone preparation system. The system is typically housed in a building. There are no specific handling requirements for the limestone or limestone slurry due to pH or other toxicity. The dust generation is minimal due to wet grinding.

6.3 Ancillary Emissions and Solid Wastes (from Consumables)

Ancillary emissions and solid waste may be generated periodically (not on a continuous basis). This may include sorbents used for gas clean-up, filter cartridges and materials used in the water treatment system, and catalyst used in reactor vessels. Spent materials will need to be replaced with fresh charge of materials as part of the plant maintenance. A more detailed discussion including the handling of the materials, their properties and the regulatory requirements for the major categories follows.

Sulfur Sorbent Media

Copper based adsorption media will be used for the desulfurization of the supplemental fuel fed to the CEPACS plant. The supplemental fuel is pipeline natural gas that is assumed to contain ~3 ppmv of sulfur odorants. Desulfurization will occur in two flow-through adsorption vessels arranged in series, each approximately 10' in diameter and containing a bed of granular media approximately 20' in height. Natural gas desulfurization adsorbents become spent when they are sufficiently loaded with adsorbates such that they can no longer remove the target sulfur compounds down to the concentration required by the process. Recent advances in desulfurization media have dramatically increased both the adsorption capacity of the media as well as the selectivity of the absorbents for the target sulfur compounds. The total spent desulfurizer media quantity over the 30-year design life of the CEPACS plant is estimated to be 2004 tons, or ~67 tons/year. It is anticipated that the spent material will be generated two times each year at ~33 tons/event.

Current experience with copper-based natural gas desulfurization adsorbents is that when they are spent they exhibit hazardous characteristics and require management under Subtitle C of the Resource Conservation and Recovery Act. This fact, together with the media's relatively low metal content, reduces the prospects of cost-effective metal recovery. Metal market prices at the time of activity, however, will dictate the eventual management method for the waste/recyclable material. Should a viable and cost-effective method of material reclaiming be available for this material stream, it will be thoroughly evaluated. Some non-target adsorbates can be captured by the media (i.e., aromatic hydrocarbon species), which could impact the waste characteristics of the spent media material.

OSHA hazcom and other General Industry Standards apply to this material/process as do EPA Tier II reporting and RCRA (Resource Conservation and Recovery Act) waste management rules and DOT Hazardous Material Rule (HMR) requirements. The desulfurizer media are not volatile, flammable or explosive. Prior to removal from the process, spent media is purged of flammable gas to render it non- flammable and safe to handle. Inert gas, nitrogen, is used to displace the natural gas from the sorbent bed.

If arrangements can be made with the utility company to provide the natural gas directly to the CEPACS plant without odorants, then the need for the desulfurization process could be eliminated or reduced, with a commensurate reduction in the generation rate and quantity of spent media.

Water Treatment System Discharge and Solid Waste

The water treatment systems employed in the plant include one for the boiler water and one for the cooling water. This subsection reviews the process and the chemicals used for the two categories of the water treatment systems. The discharge characteristics and related handling are discussed. A discussion on the solid waste related to the water treatment systems is also included.

Boiler Water Treatment and Blowdown: The control of boiler water quality includes both internal and external water treatments. External treatment includes filtration and softening to reduce the amount of particulates, chlorine, and hard minerals present in the boiler make-up water. The boiler makeup water, despite external treatment, may contain some oxygen and carbon dioxide. Since both carbon dioxide and oxygen can lead to corrosion, internal treatment with chemicals is used to reduce or eliminate these species from the system. During operation, corrosion products inside the boiler system become suspended in the water with some portion dissolving. Over time, these impurities deposit inside the pipes and on other surfaces within the system leading to poor system performance. Internal treatment includes injection of chemicals into the boiler to maintain water quality and prevent the formation of scales on the heat transfer surfaces of the boiler. Table 6-4 gives examples of chemicals generally used in the treatment of boiler water for corrosion control.

Table 6-4. Chemicals Used in Boiler Water and Cooling Water Treatments

Treatment	Chemical Most Commonly Used	Typical Concentration	EH&S Related Comments
External Water Softening	Sodium Chloride	Saturated Solution (375 g/L)	No human health concern.
Water Softening and Dispersants	Phosphates	5 – 60 ppm	Environmental impacts include rapid plant and algae growth in bodies of water leading to death of aquatic species. Minimal human safety concerns as only very high concentrations are considered dangerous.
Anti-Scaling	Lignosulphonates	1-10 ppm	Skin, eye and respiratory tract irritation, no significant environmental risk.
pH Control	KOH or NaOH	0.5 ppm	Corrosive, can cause severe burns at high concentration.
Oxygen Scavengers	Sodium Sulfite.	10-60 ppm	Minor irritation, diarrhea, stomach-ache, no long term effects. No significant environmental risk.

The boiler system is closed loop and therefore under normal operation, blowdown is the only discharge of boiler water. In the discharge of boiler water through the blowdown process, the solids and other contaminants that have built up inside of the system are removed. The blowdown water is generally sent to the site's waste treatment system for further treatment before disposal. The site's final treatment generally includes pH neutralization, biological oxygen demand control and removal of suspended solids.

Cooling Water Treatment and Blowdown/Discharge: The treatment of cooling water includes chemical injection to control corrosion within the cooling system. Table 6-4 also includes examples of chemicals generally used in the treatment of cooling water for corrosion control. An additional concern with cooling water system is that bacteria can grow due to the relatively low temperature of the cooling water. Therefore treatment programs for controlling the growth of microorganisms are employed in addition to the corrosion treatment. Table 6-5 gives the chemicals most commonly used for controlling the growth of microorganisms in cooling water.

Table 6-5. Chemicals Used for Microorganism Growth Control

Biocide Type	Examples	Typical Concentration	EH&S Related Comments
Oxidizing Biocide	Chlorine, Bromine, Chlorine Dioxide	0.5 – 1.0 ppm	Environmentally unfriendly, toxic to plants and animals, Health and Safety - irritant at low concentrations, toxic at elevated concentrations.
Non-Oxidizing Biocides	Carbamates, quaternary ammonium compounds.	~100 ppm	Depending on biocide used can be toxic to humans. Potential environmental issues.

Just as in boiler operation, cooling system must also use a blowdown procedure to remove corrosion products and the buildup of other contaminants within the cooling system. The blowdown is sent to the facility's waste treatment system for further processing before disposal. Cooling water system includes a cooling tower through which discharge of gaseous substances and water occur in addition to the normal blowdown process. The most likely chemicals discharged through the cooling tower are chlorine and bromine, especially if the pH is low.

Solid Wastes: The CEPACS plant and external boiler-water and cooling water treatment systems include water softening beds. The beds contain cation exchange media which exchange sodium or potassium ions for other cationic species, such as calcium, magnesium and iron, to reduce the scale forming potential of the feed water. The softener media are regenerated in place using concentrated sodium or potassium chloride solution and the waste stream from the regeneration process is sent to the PC plant waste treatment system. Disposal of the media is expected to occur only at plant decommissioning.

Other sources of solid waste include activated carbon beds which use activated carbon to remove organics and chlorine from the feed water, granular media filtration beds, and filter cartridges. Maintenance of equipment such as carbon and media filtration beds includes backwashing to regenerate the beds rather than replacement. Therefore minimal generation of waste carbon and filter media is expected. Filter cartridges, on the other hand, need to be replaced every 3-4 months depending on the level of particles in the water and the water demand.

Low Temperature Shift Converter Catalyst

The Low Temperature Shift Converter (LTSC) Catalyst is required to shift the CO present in pre-cooled anode exhaust stream to convert it to CO_2 for capture and produce additional hydrogen. It is a copper based catalyst containing 40-60% CuO (32-48% copper). The total quantity of LTSC Catalyst is estimated to be 72 tons. It is expected that the initial LTSC Catalyst charge will operate through the design life of the CEPACS plant. Also considering the clean conditions of exposure and the relatively high metal content of the catalyst, it is expected that spent catalyst material can be recycled to reclaim the metal content.

OSHA hazcom and other General Industry Standards apply to this material/process as do EPA Tier II reporting and RCRA waste management rules and DOT Hazardous Material Rule (HMR) requirements. The LTSC Catalyst is not anticipated to be volatile, flammable or explosive. The spent LTSC catalyst must be conditioned prior to discharge to prevent excessive catalyst heating upon exposure to air. Inert gas is used to displace shift gas, and prevent oxygen intrusion into the catalyst bed. Catalyst vendors provide detailed guidance on catalyst deactivation procedures and professional firms are available to perform the procedure.

Following deactivation/conditioning, the spent LTSC catalyst can be safely removed and handled in accordance with routine methods.

ECM Stacks

The CEPACS plant uses 2000 ECM stacks – 200 ECM stacks in each of 10 ECM module enclosures. The CEPACS plant has a design life of 30 years and the ECM stacks have a design life of 10 years. This requires that all the ECM stacks be replaced twice. The quantity of ECM stacks to be used over the life of the CEPACS plant will be 6000.

The material related composition of the ECM stacks is included in Table 6-6.

Table 6-6. Material Related Composition of ECM Stack

Material Class	Composition (%)	Weight (lb)
Stainless Steel	16.7	3010
Cell Packages	82.1	14772
Exotic Metals	0.4	76
Dielectrics	0.8	142
Total	100.0	18,000

The disposition for each material class is as follows.
<u>Stainless steel</u>: About 74% of the stainless steel parts from each spent ECM stack (2237 lb.) are candidates for immediate reconditioning and reuse in replacement ECM stacks. Any parts that are unsuitable for reconditioning and reuse are 100% recyclable as high value stainless scrap metal. All of the remaining parts are 100% recyclable.
<u>Exotic Metals</u>: The exotic metal components are 100% recyclable as high value scrap metal.
<u>Dielectrics</u>: The dielectric components are all candidates for reconditioning and reuse. Broken or unusable dielectric components are disposed of.
<u>Cell Packages</u>: Cell packages are 100% recyclable. Cell package components are high in nickel and chromium content and, in the proper form, are valuable in the secondary metals market. Under carefully managed conditions, the spent ECM stacks can be disassembled into forms that are capable of being converted (melted down) into metal ‘pigs’ that can be used as ingredients in stainless steel manufacturing. Materials managed in this manner can be excluded from the definition of hazardous waste, or at best may even be excluded from the definition of solid waste. Due to high nickel and chromium content of the cell package components and the presence of some refractory material, the cell packages have a very high melting point, requiring the addition of ferrous metal ‘bushing’ into the foundry crucible during the melt process. The results of a trial melt of a spent fuel cell (ECM) stack, including added ferrous bushing, are provided in Table 6-7.

Table 6-7. Material Composition of Trial Melt of Used ECM Stack

Element	Composition (%)	Within Spec
Ni	27.5	Yes
Cr	6.3	Yes
Fe	63.6	Yes
Minor metals	< 2.0% (total)	Yes
Non-metals	< 0.8% (total)	Yes

Metal pigs of the above formulation are marketable in the secondary metals market. In summary, the material classes of the spent ECM stacks will be managed as shown in Table 6-8.

Table 6-8. Material Related Management of Used ECM Stack

Material Class	Composition (%)	% of material class reused	% of material class recycled	% of material class discarded
Stainless Steel	16.7	0-74	26 – 100	0
Cell Packages	82.1	0	100	0
Exotic Metals	0.4	0	100	0
Dielectrics	0.8	0-100	0	0 – 100

The primary environmental, health and safety regulations that impact the spent ECM stack management programs are the EPA RCRA (Solid and Hazardous Waste regulations) and the OSHA General Industry Standards. RCRA regulates the correct management of spent material with emphasis on materials recovery versus waste disposal. The materials that comprise the ECM stacks are prime candidates for materials recovery. With the prudent and proper application of the RCRA regulations, very high rates of materials recovery can be realized.

6.4 Incidental Emissions/Releases

The CEPACS plant includes chillers used for post-processing of the CO_2-rich anode exhaust stream from ECM stacks. The cooled shifted anode gas stream is chilled and compressed in stages to liquefy CO_2 before flash separation to near pure CO_2 product stream. The chiller units use a closed-loop of ammonia or other refrigerant. Also, the CEPACS plant includes a Polishing FGD unit for removal of SO_x from the flue gas to levels required for suitable operation of ECM stacks. Lime slurry is prepared and fed to the FGD unit and gypsum is produced as a byproduct. Incidental emission or release of ammonia (refrigerant) and lime or gypsum-related emissions are reviewed in this section. Information on handling, properties and regulatory requirements is also included in the subsections that follow.

Ammonia or Other Refrigerants

Ammonia (NH_3) is a refrigerant (R-717) required by the CEPACS system. The CEPACS process uses two stages of absorption chillers to condense the CO_2. The ammonia is contained in a closed loop that circulates through the chiller units. Discussions with vendors indicated that the technology is reliable and essentially leak-free. The estimated total charge of ammonia in the chillers to provide the required cooling capacity for the CEPACS process is 36 tons. It is estimated that the initial refrigerant fill will be used for the design life of the plant.

At atmospheric conditions, ammonia is toxic and potentially flammable gas. It poses an inhalation hazard at concentrations above 25 – 50 parts per million (ppm) and is a respiratory irritant. It is considered Immediately Dangerous to Life and Health at concentrations above 500 ppm. Ammonia is flammable at concentrations in the range of 16- 25% in air. It is highly soluble in water.

Ammonia is a commonly used industrial refrigerant gas. Industrial processes that contain more than 10,000 pounds of ammonia are subject to the provisions of the Occupational Safety and Health Administration's general industry standard on Process Safety Management (PSM, 29 CFR 1910.119) as well as the EPA's regulation on Risk Management Planning (RMP, 40 CFR Part 68) on Chemical Accident Prevention Provisions. The two regulatory standards are closely aligned in scope, so the mandated compliance requirements are normally coordinated into a common compliance program. The host PC plant for the CEPACS plant is likely to have SCR

NOx removal, and therefore also likely to employ either ammonia or another PSM/RMP chemical required for SCR injection and catalytic reaction. PC plant SCR NH_3 injection would be approximately 400 lb/h, or approx. 1400 ton/yr of NH_3 consumption. Ammonia or ammonia-containing chemical or solution use at this scale would very likely trigger the provisions of the PSM and RMP. It is likely that the additional ammonia contained in the CEPACS absorption chiller units would represent a modification to an existing PSM/RMP program plan.

Lime, Gypsum

Limestone ($CaCO_3$) is used in the CEPACS polishing FGD unit at a rate of 868 lb/h. The polishing FGD produces 1,351 lb/h of $CaSO_4$ (gypsum) adding to the PC plant solids dewatering plant load. These mass flow rates are incremental rates associated with the polishing FGD only and represent ~2% increase in the overall mass flow rates for the respective streams for the host PC plant. The limestone and gypsum handling processes and equipment for the CEPACS plant polishing FGD are essentially identical to those used by the PC plant primary FGD. It is assumed that there exists adequate capacity in the limestone slurry and gypsum dewatering equipment to handle these increases in throughput.

Limestone is typically >98.5% calcium carbonate. $CaCO_3$ is a naturally occurring non-toxic earthen material that is nearly innocuous. It is non-flammable and non-reactive. In its dry form, it can produce dust that under severe, long-term human exposure conditions can result in pneumoconiosis (lung disease). Limestone can also contain trace concentrations of other minerals, including silica (quartz.) Respirable crystalline silica is a suspected carcinogen. The limestone processes for both the PC plant and CEPACS system are wet slurry processes, so dust exposure concerns are manageable. Control equipment will be used to limit dust generation and exposures to the incoming limestone unloading operations.

Calcium sulfate, also known as gypsum, is a non-flammable, non-reactive, non-volatile solid and non-toxic compound. In commercial forms (wall board), it may contain small quantities of crystalline silica. But as produced from the dewatering of FGD blowdown, it will only contain as much silica as was introduced through the limestone feed. Gypsum can generate dust, but as mentioned earlier, the FGD processes are wet processes and dust generation should not be a problem.

7.0 BENCH-SCALE TESTING OF ECM-BASED CO_2 CAPTURE SYSTEM

7.1 System Design and Construction

A demonstration system was designed for bench-scale testing of ECM to prove the feasibility of using ECM technology for CO_2 capture. The system utilized an ECM stack containing cells with a total electrochemical membrane area of 11.7 m^2, expected to produce a gross DC output of 9 kW. The bench-scale system was implemented via modification of one of FCE's existing test facilities. The facility included the system key balance-of-plant equipment demonstrating the overall capabilities of the CEPACS system. Both the anode and cathode (ECM) feed streams were a mix of different gases to simulate what the anode and cathode streams were expected to be in the CEPACS system. When the vented CO_2 (in the facility system) was accounted for, the expected CO_2 captured from the simulated flue gas was ~91%. This met the requirement of capturing at least 90% of CO_2 from flue gas. The process flow diagram for the demonstration system was developed. Design of the system was guided by the existing FCE test facility selected and the type of equipment available. System simulations were carried out using CHEMCAD process design software. Table 7-1 presents a summary of the simulation results for the demonstration system.

Table 7-1. Summary of Demonstration System (10 kW CEPACS) Simulation Results

10 kW CEPACS SYSTEM SUMMARY	PARAMETER		ENGLISH	UNIT		SI	UNIT
	CARBON DIOXIDE CAPTURE OVERVIEW						
	Amount of CO2 Captured		0.27	tons/day		0.25	tonnes/day
			99.90	tons/year		90.63	tonnes/year
	Net Coal Plant CO2 Captured		0.18	tons/day		0.16	tonnes/day
			65.00	tons/year		58.97	tonnes/year
	Percentage of Coal Plant CO2 Captured		90.1	%			
	Total Percentage of Carbon Captured		93.3	%			
	SUPERCRITICAL CARBON DIOXIDE PRODUCT INFORMATION						
	Pressure		2214.7	psia		152.7	bara
	Vapor Fraction		0				
	Mass Flow Rate		22.9	lb/hr		10.4	kg/hr
	Carbon Dioxide Purity		99.3	%			
	FUEL CELL POWER PLANT OVERVIEW						
	Cell Voltage					0.7400	V
	Current Density					110.0	mA/cm^2
	Carbon Dioxide Flux					135.0	cc/m^2-s
	FUEL UTILIZATION		75%				
	ECM Operating Temperature		1126.6	°F		608.1	°C
	ECM MODULE OVERVIEW						
	Number of ECM Cells					15.0	
	Cell Active Area					7825.0	cm^2
	Total ECM Area					11.7	m^2
	ECM Module DC Output					9.6	kW
	Efficiency, HHV					46.1	%
	Natural Gas Consumption		1.10	scfm		1.87	scmh
	DI Water Supplied		0.04	gpm		0.15	Lpm
	Process Water Supplied		2.25	gpm		8.51	Lpm

Process data sheets to supplement the equipment specifications were prepared for all major equipment based on system simulation results. The process specifications for the CO_2 (product stream) compressor and ECM (stack) module are presented in Tables 7-2 and 7-3, respectively, as examples.

Table 7-2. Compressor Process Specifications

COMPRESSOR		
400-CR-100		
	Inlet	Outlet
Temperature, °F	70.00	563.89
Pressure, psia	15.00	250.00
Flow Rate, lbmol/hr	0.84	0.84
Flow Rate, lb/hr	32.62	32.62
Composition		
Hydrogen	0.0724	0.0724
Water	0.0243	0.0243
Methane	0.0443	0.0443
Carbon Monoxide	0.0058	0.0058
Carbon Dioxide	0.8498	0.8498
Oxygen	0.0000	0.0000
Nitrogen	0.0033	0.0033
Argon	0.0000	0.0000
Ethane	0.0000	0.0000
Propane	0.0000	0.0000
Butane	0.0000	0.0000

Table 7-3. ECM Module Process Specifications

ECM MODULE				
200-ECM-100				
	Anode Inlet	Anode Outlet	Cathode Inlet	Cathode Outlet
Temperature, °F	1100.00	1071.09	1024.38	1081.09
Pressure, psia	16.05	15.95	15.65	15.55
Flow Rate, lbmol/hr	0.78	1.47	5.78	4.98
Flow Rate, lb/hr	10.93	42.66	165.87	134.15
Composition				
Hydrogen	0.4662	0.0905	0.0000	0.0000
Water	0.2421	0.3938	0.1268	0.1469
Methane	0.1018	0.0006	0.0000	0.0000
Carbon Monoxide	0.0772	0.0529	0.0000	0.0000
Carbon Dioxide	0.1091	0.4604	0.0995	0.0092
Oxygen	0.0000	0.0000	0.0827	0.0428
Nitrogen	0.0035	0.0019	0.6874	0.7968
Argon	0.0000	0.0000	0.0037	0.0043
Ethane	0.0000	0.0000	0.0000	0.0000
Propane	0.0000	0.0000	0.0000	0.0000
Butane	0.0000	0.0000	0.0000	0.0000

Equipment selection and procurement was carried out through vendor contacts. Some of the equipment was designed and fabricated in-house. Line sizing for the piping was based on flow rates calculated by the process simulation runs. Stainless steel tubing was chosen. Valve sizing was based on identified process requirements.

The selected test facility was modified to accommodate the CEPACS system tests. A three-dimensional model was developed to detail the general arrangement of the equipment in the test facility. Required tie-in points to the existing facility were identified. The facility build consisted of two major components: existing facility modifications and CEPACS skid construction. A Hazard and Operability (HAZOP) Study was performed on the existing facility to ensure that it was properly designed to handle the modified process as expected. A few modifications deemed necessary per the HAZOP were implemented.

Construction of the CEPACS skid was completed. A picture of the skid is shown in Figure 7-1. The skid was constructed utilizing a Unistrut base and superstructure to support the process equipment. The skid was built per detailed fabrication drawings that were created from a 3-dimensional CAD model. All process equipment was fabricated/procured and installed, including: Low Temperature Shift Converter, water separators with level controllers, CO_2 compressor, CO_2 condensing heat exchangers, liquid CO_2 separator, and a chiller capable of supplying -76°F coolant.

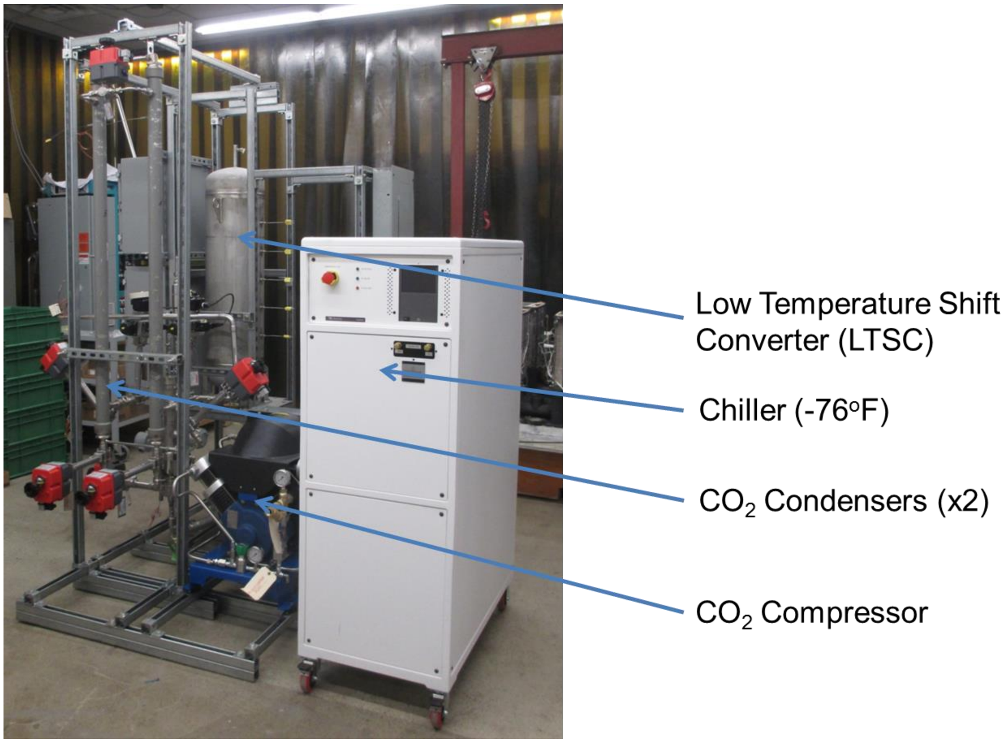


Figure 7-1. Picture of CEPACS System Skid for Post-Processing ECM Anode Exhaust into High-Purity Liquid CO_2 Product

Fabrication of the 11.7 m^2 ECM test stack, including the associated module enclosure hardware was also completed. Figure 7-2 shows pictures of the completed stack assembly, prior to addition of gas manifolds and enclosure (left), and after installation of the gas manifolds and vessel base insulation (right). The test stack included 14 full-area cells. The cell assemblies were obtained from FCE's Torrington, CT commercial fuel cell manufacturing plant. The cells were assembled into a stack which was then installed into an ECM module at FCE's Danbury, CT R&D headquarters. The stack hardware (end-plates, compression system, manifold design, etc.) was based on the designs used in full height (401-cell) commercial fuel cell stacks.

Figure 7-2. Pictures of 14-Cell ECM Test Stack: Before (left) and after (right) installation of gas manifolds and insulation

The CEPACS anode-exhaust processing skid was integrated into the test facility along with the 14-cell ECM stack module. Piping, power related wiring and instrumentation wiring interconnections were completed. The skid was then insulated to conserve heat (during operation) in both the warm (~500°F) and cold (-50°F) process sections. The system was made ready to commence testing in BP3 of the project.

7.2 Demonstration Testing

Bench-scale testing of an 11.7 m^2 ECM-based CO_2 capture system was conducted during BP3 of the project. After conditioning (initial start-up) of the stack, the stack baseline electrochemical performance was validated at standard DFC (Direct Fuel Cell) operating conditions. The ECM stack was then operated at CEPACS system operating conditions using simulated PC plant flue gas. Figure 7-3 shows the test results during the total test period of >15,700 h which included steady state testing as well as the parametric and optimization tests. The stack completed the planned nine months of steady state testing, meeting one of the project milestones. The CO_2 flux of 116 cc/s/m^2 was maintained constant for over 6,500 hours, transferring more than 90% of the CO_2 in the cathode stream to the anode stream. The project technical milestone of verifying CO_2 flux at 100 cc/s/m^2 was achieved, as the CO_2 flux observed in the bench-scale test exceeded the required flux. Additionally, the ECM stack had a constant gross DC power output of nearly 8 kW (>10 kW peak power), and experienced a power degradation rate of 0.05%/1000 h. The power degradation rate was less than one third of that observed for FCE's sub-scale fuel cell stacks operating in power generation mode, likely due to reduced in-stack thermal gradients observed when operating in carbon-capture mode. As shown in Figure 7-4, the stack transferred more than 120 tonnes (metric tons) of CO_2 from the simulated flue gas to the anode exhaust stream over the total test period, while simultaneously generating more than 110 MWh of gross DC electricity.

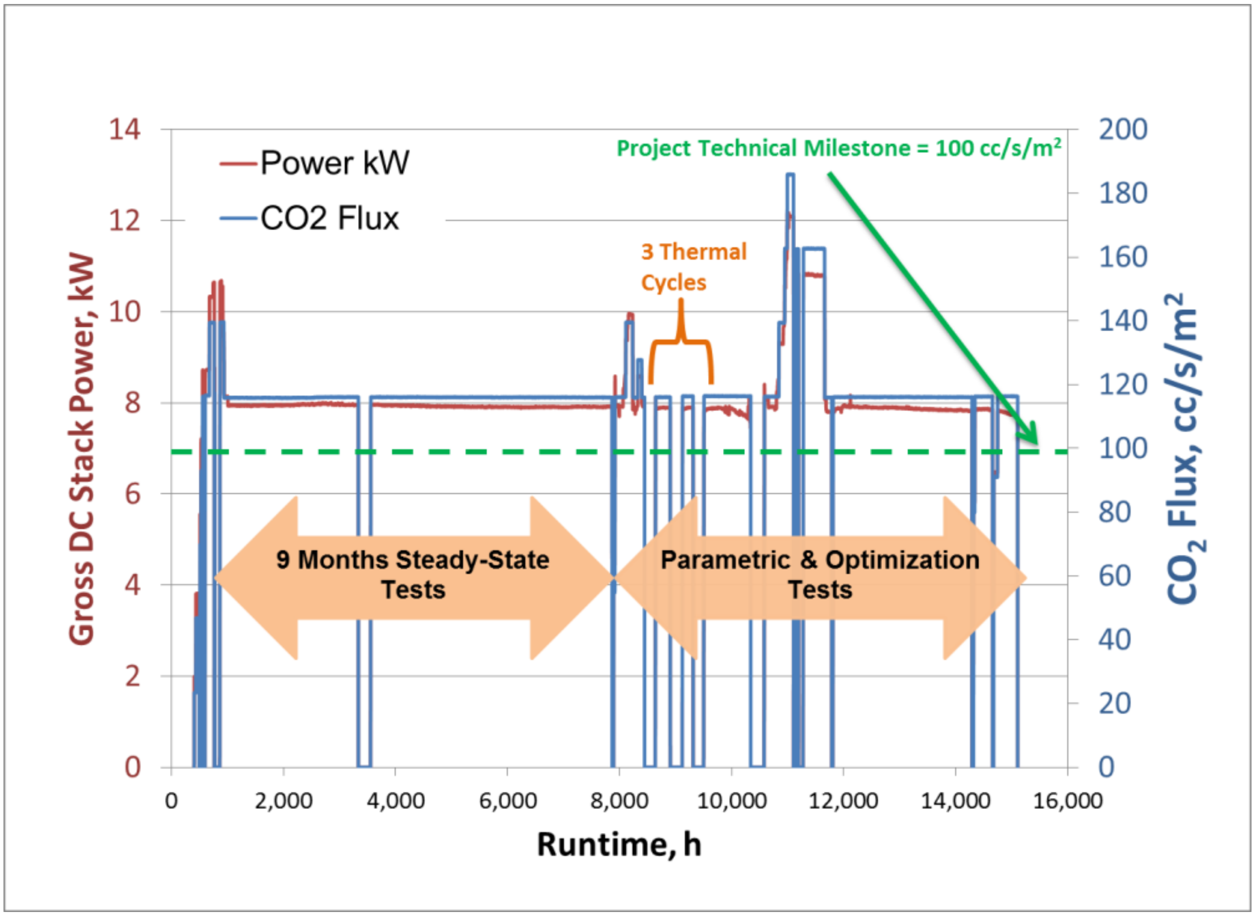


Figure 7-3. Carbon Capture and Power Generation Performance of 14-Cell ECM Stack during Long-term Steady State and Other Tests

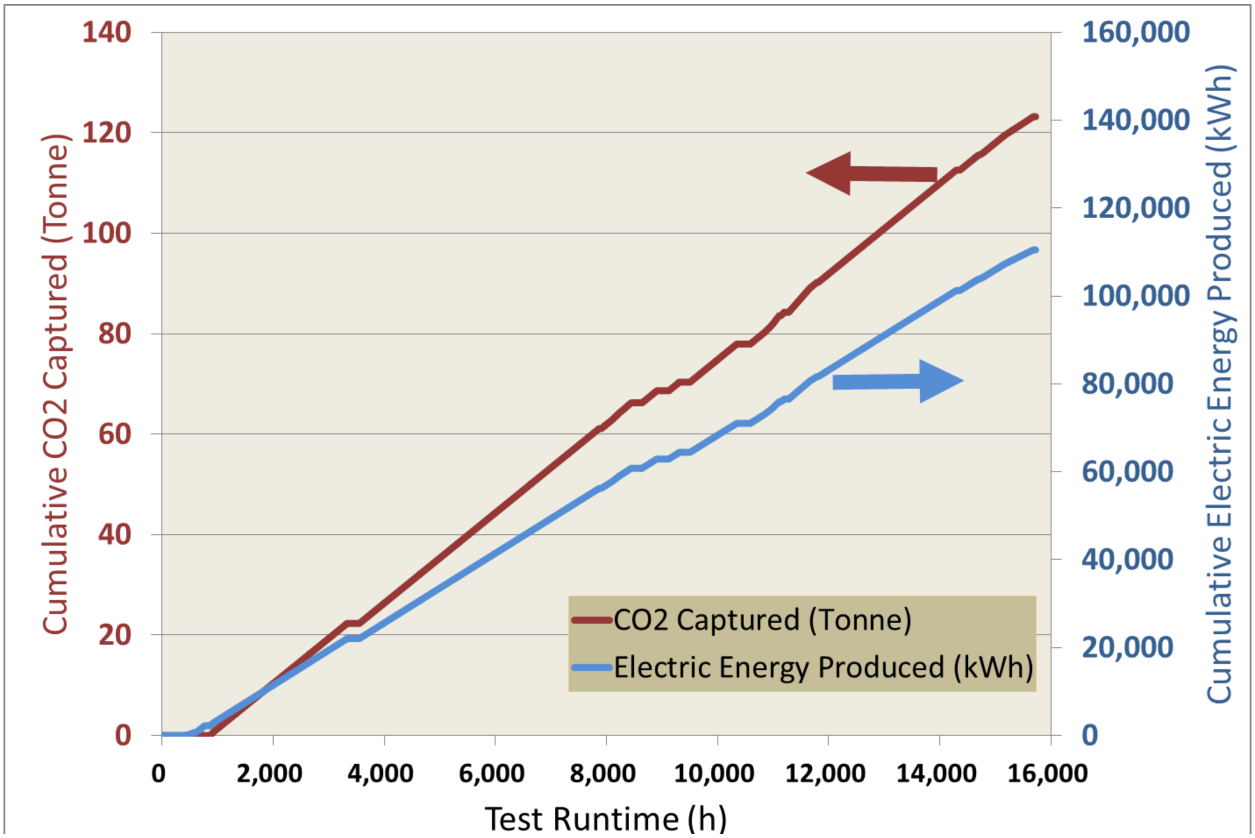


Figure 7-4. Cumulative CO_2 Transferred and DC Electrical Energy Generated by 14-Cell ECM Stack during Bench-scale Testing

Parametric testing, including operation at various carbon capture percentages and current densities (CO_2 flux), was conducted. Figure 7-5 shows the test results. The testing showed that the ECM stack is capable of operating at 20% higher CO_2 flux (~140 cc/s/m^2), with a proportional increase in power output. As the carbon capture percentage was decreased, the ECM stack voltage increased corresponding to higher system efficiency. Parametric Testing was also performed to optimize supplemental air addition for PC plant flue gas. As shown in Figure 7-6, the O_2 content of 10% (slightly higher than that observed in single-cell testing) was found to be optimal.

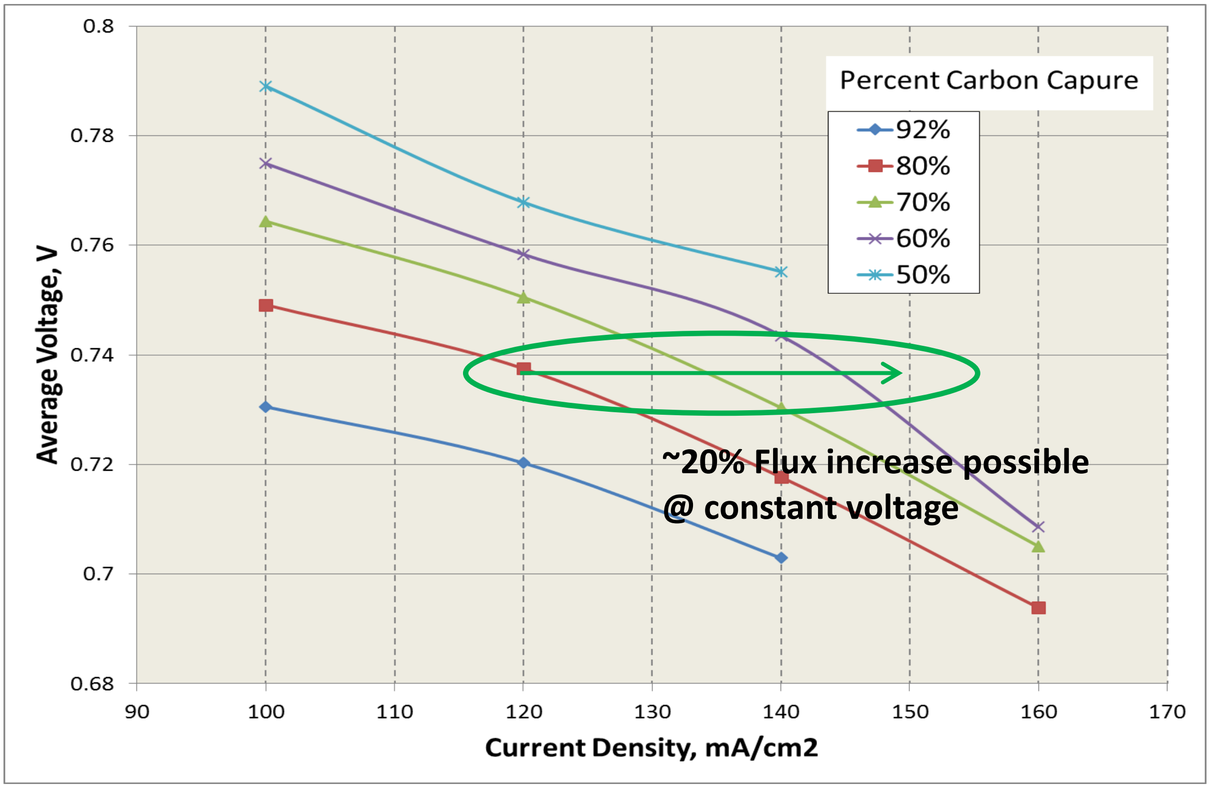


Figure 7-5. ECM Stack Performance Characterization at Various Carbon Capture Levels during Bench-scale Testing

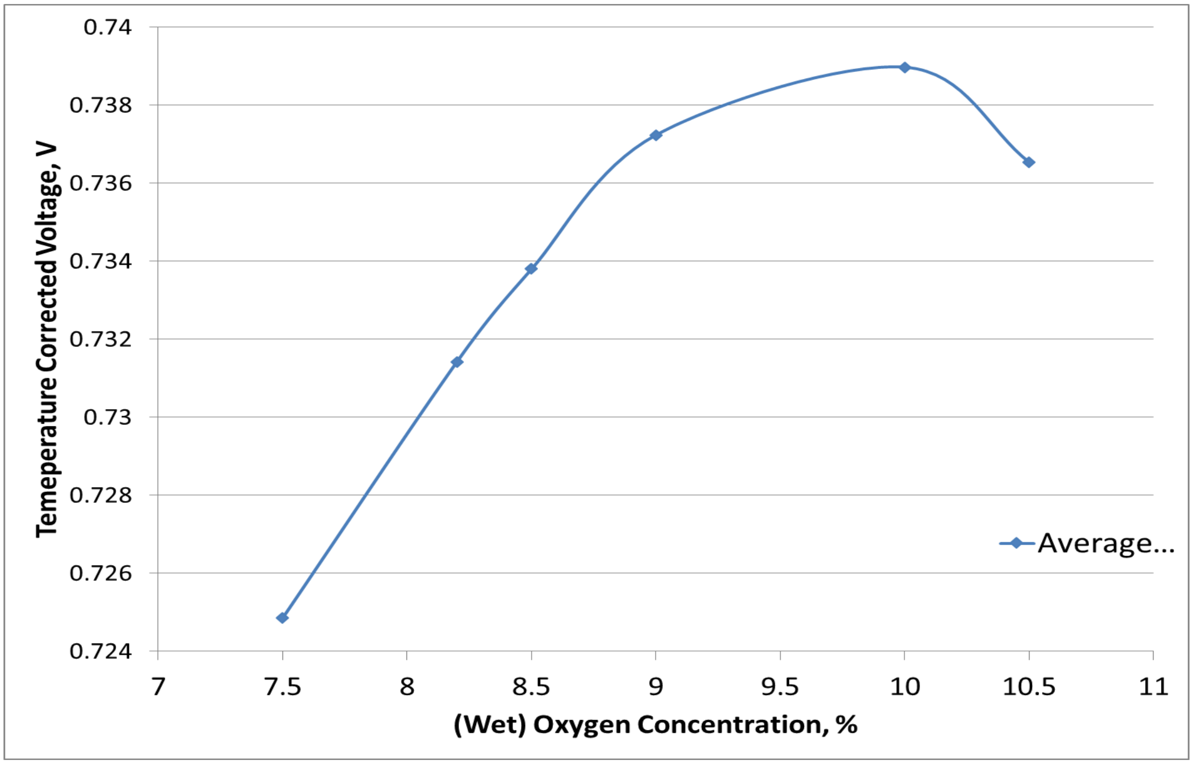


Figure 7-6. Optimization of ECM Cathode Feed O_2 Concentration

Other parameters studied during the parametric tests included stack temperature and fuel feed composition. Thermal cycling of the ECM stack was also performed. The stack was cooled from around 600°C (operating temperature) to <100°C. Following each thermal cycle, there was an extended period of operation at the steady state hold conditions. The required three thermal cycles were completed. Figure 7-7. shows the stack temperatures during these thermal cycle operations. Figure 7-8. shows the stack performance (individual cell voltages) during the steady state hold period after each thermal cycle. There was a minimal performance penalty, in terms of DC power produced and no performance penalty in terms of CO_2 flux or Current during the thermal cycling evaluation.

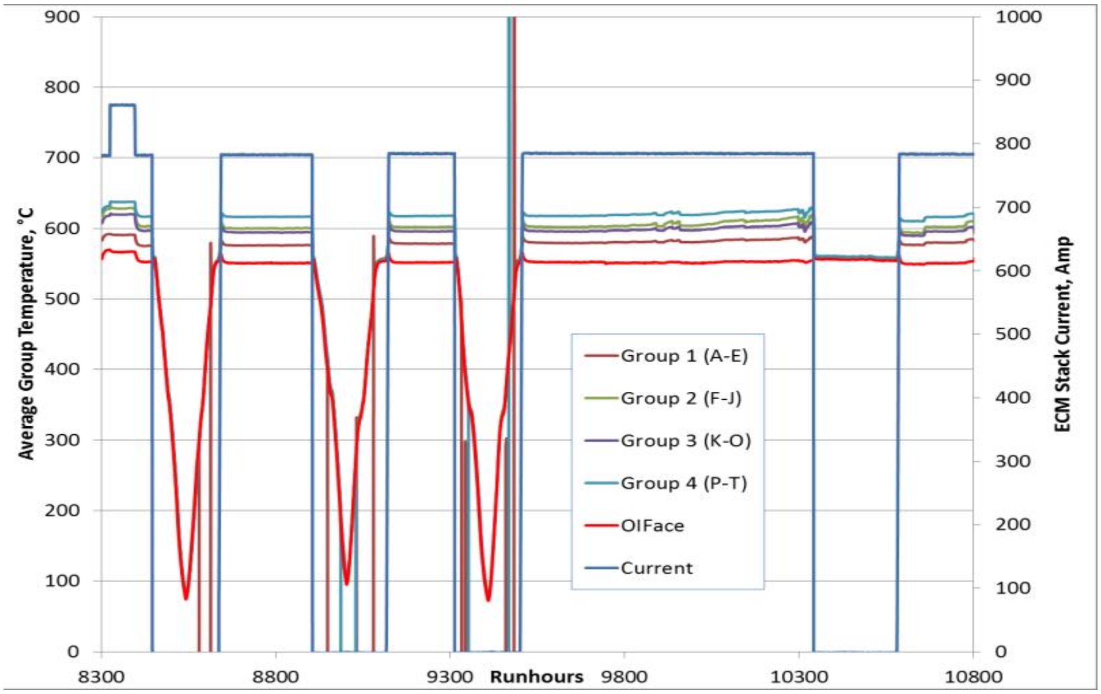


Figure 7-7. Average Cell Group Temperatures during Thermal Cycling of 14-cell ECM Stack

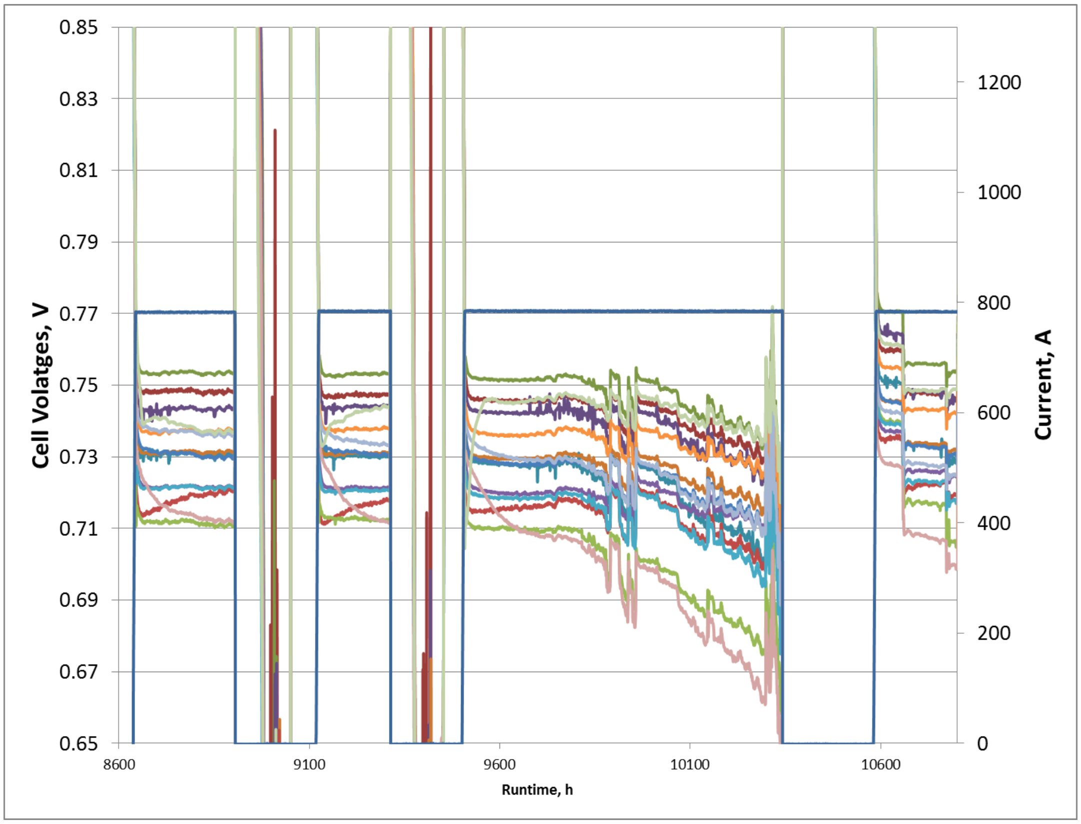


Figure 7-8. Individual Cell Voltages during Thermal Cycling of 14-cell ECM Stack

As mentioned earlier, the bench-scale test facility included a CO_2 compression/liquefaction skid (representing the part of the CEPACS system that post-processes the anode exhaust stream to produce high-purity liquid CO_2 product). Testing of the CO_2 compression/liquefaction skid was carried out in parallel (with the bench-scale ECM stack CO_2 capture tests). After the CEPACS CO_2 capture skid was installed in the (10 kW) test facility, mechanical checkouts were performed on all major equipment, including the low temperature shift reactor (LTSR) heaters, compressor, water pump and chiller. Process control checkout testing of the skid was also conducted. The low temperature shift reactor catalyst was reduced and the skid was evaluated under simulated ECM stack exhaust conditions, demonstrating the effective liquefaction and separation of high-purity CO_2. Additional testing included optimizing the operation of key system components. Several major advancements occurred in both the physical system hardware and system testing. To effectively evaluate operability at a wide range of operating conditions, system reliability upgrades were completed and any system deficiencies limiting extended duration testing were addressed.

Following the system upgrades, a variety of operating conditions were applied to the CO_2 purification skid to characterize the system response, product characteristics and validate the operation. It became clear during extended testing that there were hardware limitations which rendered consistent, repeatable results difficult to obtain. A stepwise investigation of the problem indicated two major issues which were addressed sequentially. Pure CO_2 product was delivered via a simple drop pot mechanism utilizing two level sensors in the liquid CO_2 separator to ensure that only CO_2 was exhausted. However, the vaporization system designed to process this liquid CO_2 consistently clogged resulting in system overpressures and inconsistencies. Further investigation revealed an insufficient margin within the vaporizer heating system and several damaged components. The underlying problem was traced to variability within the dropout system whereby certain liquid CO_2 dump events exhibited substantial variation in delivered liquid CO_2. During normal operation, the system was capable of processing all delivered CO_2. However, during these high volume dump exceptions, the heating system was unable to cope with this excess CO_2 resulting in regulator freeze-up and damage. By adding a secondary regulator to reduce the depressurization endotherm and increasing the vaporization heater margin, it was possible to extend operation dramatically.

During the investigation of the vaporization system, it was also discovered that certain system temperature and pressure set points would result in CO_2 separator clogs even with the upgraded vaporization system. Careful monitoring of the CO_2 liquid level sensors during steady state operation revealed a steadily decreasing dump frequency resulting in the eventual freeze up of the vaporization system or a separator overfill event. Further exploration of the liquid level sensors indicated the potential for electronics failure due to under-temperature events. During short term or summer operation with high ambient temperatures, these sensors functioned well. However, extended runs in the fall and winter with lower ambient temperatures resulted in electronics errors (not correctly reporting level), leading to excessive volume CO_2 dumps or CO_2 separator overfill events. By thermally isolating the sensor electronics package and increasing ambient airflow, it was possible to increase the sensor reliability, enabling unattended and extended operation.

The CO_2 capture skid was tested over a relatively wide range of operating parameters. The results in general indicated the ability to produce food grade liquid CO_2 (>99.9% CO_2 and no measureable CO) across all tested operable points. Table 7-4 lists one set of test conditions as an example.

Table 7-4. Example of Input Conditions Used During CO_2 Purification Tests

Input Conditions – Realistic Testing	
Nitrogen Flow	0.0 SCFM
CO_2 Flow	3.7 SCFM
Hydrogen Flow	0.9 SCFM
Water flow	47 g/min
Pressure	250 PSI
Temperature	-53°C

System performance was assessed by gas chromatography (GC) analysis at three points: downstream of the LTSR, pure CO_2 stream line, and system exhaust line. In an integrated system, the LTSR would convert excess CO into CO_2. However, as no CO was injected during this testing, CO levels registered near equilibrium levels as expected and provided the expected composition and flow of gas into the CO_2 separation system. As noted previously, the pure CO_2 GC analysis consistently indicated 99.9% CO_2 with trace amounts of hydrogen contamination. By performing mass balance analysis based on the inlet and exit gas flow rates and compositions, it was possible to determine capture rates. CO_2 separator temperature appeared to be the dominant system variable controlling the variation in rate.

Testing was transitioned from system development stage to system prototype maturation, operator training and system optimization for long term data collection processes. Further optimization was made related to the CO_2 capture and purification skid to enable easier operation for extended duration testing and more reliable data collection. A key outcome was the training and validation of several system operators who were not involved with the development of the system hardware. This transition demonstrated the maturity of the CO_2 capture and purification system and its ability to be used as a parametric analysis tool rather than a development prototype. Operators participated in a three day intensive training course covering system operating principles, human interface, development history, hardware/software components and hands-on operational experience. Following the training course; operators successfully demonstrated system startup, CO_2 separation and purification validated by GC sampling and system shutdown with no developer interactions. The parametric system data for a wide range of system operating points will be used to validate system models of large scale carbon capture systems and more fully elucidate desirable operating points.

8.0 UPDATED TECHNICAL AND ECONOMIC FEASIBILITY STUDY

The final updated Technical & Economic Feasibility Study was conducted in Budget Period 3 of the project to assess the performance and cost of FCE's ECM-based CO_2 capture system. The CEPACS plant utilizing ECM technology was designed to capture and compress >90% of CO_2 from the flue gas of a reference 550 MW (net AC) Pulverized Coal (PC) Subcritical Steam Cycle Plant. The PC plant design specified in "Case 9" of the U.S. Department of Energy - National Energy Technology Laboratory (DOE - NETL) report "*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2a"* [26] was used as a reference.

The system configuration, simulations and analyses were performed using CHEMCAD process simulation software to guide the conceptual design of the CEPACS plant. Process simulations were updated based on ECM performance realized in bench-scale testing. Average cell voltage as a function of the CO_2 transfer rate (or flux) in a carbon dioxide utilization range of 50 to 92% was shown in Figure 7-5 (presented earlier in Section 7.2). At the design condition of 128 scc/s/m², the cell voltage is ~725 mV. Adjusting for cell performance degradation, this cell voltage is equivalent to 730 mV, which was used as the basis for the design of the CEPACS system.

In the CEPACS plant, ECM-separated CO_2-rich stream is cooled, compressed and chilled to liquefy CO_2 which can be easily pressurized for sequestration or beneficial use. Technical information provided by leading balance-of-plant (BOP) equipment manufacturers was utilized for system analyses. The performance assessment included estimation of the parasitic power consumption for >90% CO_2 capture and compression, and the efficiency impact on the PC plant. The updated CEPACS plant contains a total of 1,792 ECM stacks. An equipment list, ECM stacks packaging design, and CEPACS plant layout were developed to facilitate the economic analysis. Technical and cost quotations were solicited from leading vendors for key pieces of equipment. In addition to the process flow diagram and a stream table, electrical one-line diagrams were generated for the CEPACS plant. The economic feasibility study included estimation of CEPACS plant capital cost, cost of electricity (COE) analyses and, estimation of cost per ton of CO_2 captured and cost per ton of CO_2 avoided. The study focused on developing technical and economic comparisons to Cases 9 and 10 related to Subcritical PC plants [26]. The ECM stack module enclosure concept and module layout for the plant are discussed in detail in Section 8.1. The results of the system and economic analyses are summarized in Section 8.2.

8.1 ECM Stack Module Enclosures and Layout

ECM Enclosures: The CEPACS system for separating CO_2 from the flue gas of a 550 MW PC power plant employs 1,792 ECM stacks divided into 8 sections. For large-scale applications of the CEPACS system, grouping the stacks into larger enclosures reduces the overall capital costs by eliminating smaller module enclosures and their associated piping, facilitates the replacement of individual stacks (compared to replacing complete modules) and provides economies of scale that are not possible with 448 separate 1 MW (nominal), 4-stack modules (BP1 stack packaging concept).

In each section, 224 ECM stacks are located in an enclosure. The enclosure is 70’ wide, 178’ long and 50’ high, sitting on a concrete pad 108’ wide and 181’ long. The ECM stacks are located in 8 rows with 28 ECM stacks per row as shown in Figure 8-1. The ECM stacks are supported in the enclosure on concrete and stainless steel supports. The enclosure has concrete walls and steel trusses mounted above the ECM stacks. Thermal insulation supported by the trusses, and on the walls and lower floor of the enclosure, completely surrounds the 224 stacks. A cross section of the enclosure is shown in Figure 8-2.

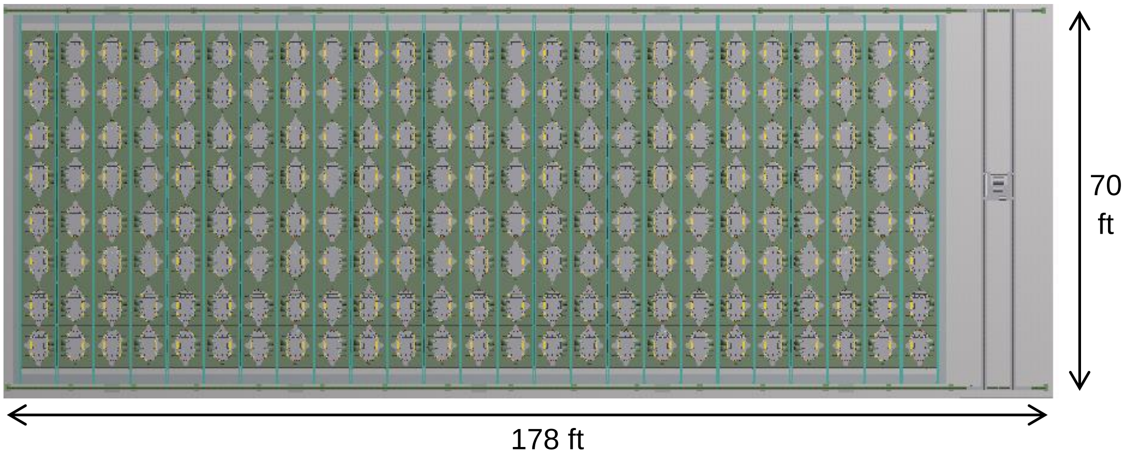


Figure 8-1. Plan View of a 224-Stack ECM Enclosure

A region in the lower part of the building is open for PC plant flue gas and air to enter the building and distribute upward to each of the ECM stack cathodes. The cathode exhaust is collected in eight 30” x 50” ducts, located below the ECM stacks, and flows to the end of the enclosure where it is collected and flows to fuel superheaters located outside the enclosure. Fuel gas from the superheaters enters the enclosure and is distributed in four 26”x 18” ducts in the lower part of the enclosure. The fuel gas flows upward from the ducts, in distributor tubes, to the ECM stack anodes. The CO_2-rich anode exhaust gas, carrying the CO_2 separated from the ECM cathodes, exits the ECM stacks and flows downward in collector tubes to two 24” x 17” ducts and three 32” x 28” ducts located in the lower part of the enclosure. The anode exhaust gas flows in the five ducts to the end of the enclosure where it is collected and routed outside the enclosure. The ducting for gas distribution in the enclosure is made from 316SS, and has expansion joints to accommodate thermal expansion.

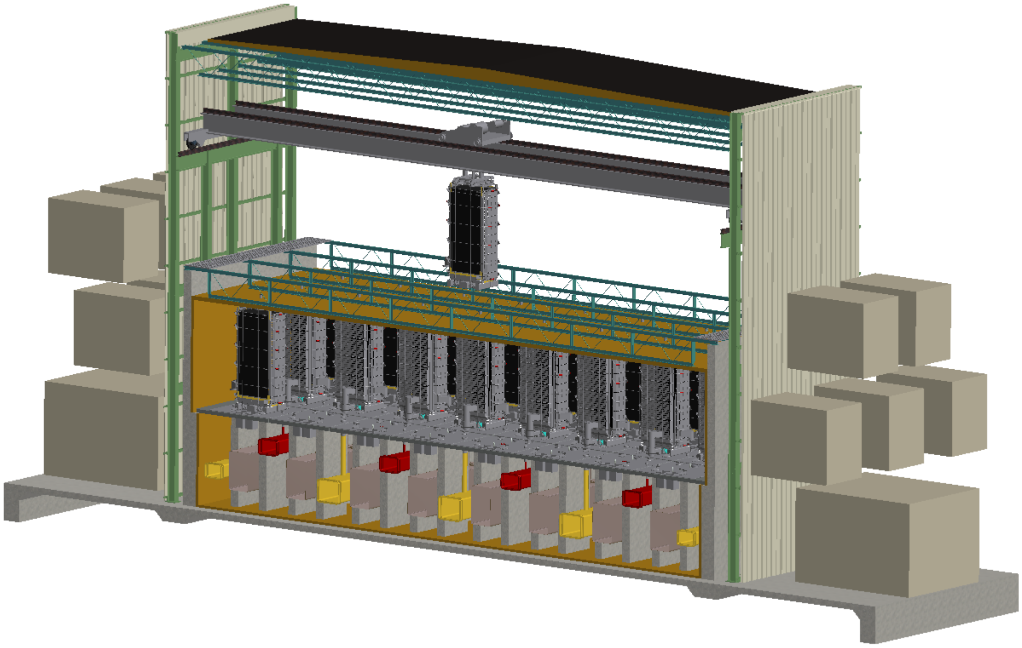


Figure 8-2. Cross-section of an ECM enclosure with stack being installed and power equipment external to the enclosure

The space above the trusses is a cool region. Bus-bars that transfer the DC current from the ECM stacks to the power conditioning equipment are located in this cool region. The power conditioning choppers and inverters are located outside and adjacent to the enclosure. Access to each of the ECM stacks is provided by removable sections of the upper insulation. A traveling crane is located in the building for placement of and access to each of the ECM stacks. The enclosure has sheet metal siding and sheet metal roofing supported on roof trusses.

CEPACS Plant Layout: The layout of the CEPACS system was designed to minimize capital costs. Specifically, the “hot” balance-of-plant equipment was de-centralized into 8 separate sections with one section located proximate to each of the 8 ECM enclosures as shown in Figure 8-3. These “hot” sections comprise such major BOP equipment as the flue gas blower, the flue gas heater, the boiler, the fuel preheater, the air blower, the air preheater, the duct burner and oxidizers, the anode exhaust cooler and the low-temperature shift converter. This modular design minimizes the lengths of “hot” piping and the quantities of fittings, significantly reducing capital costs while simplifying the sparing of parts and potentially increasing the capacity factor.

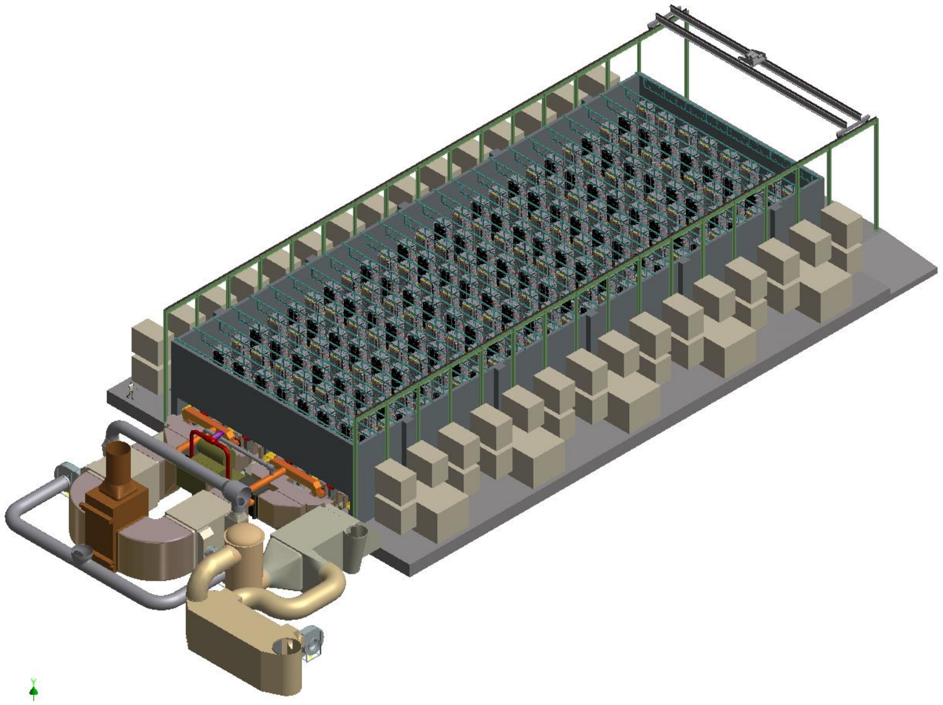


Figure 8-3. General Arrangement of an ECM enclosure and associated "hot" BOP equipment

The general arrangement of the ECM enclosures and associated "hot" BOP equipment is shown in Figure 8-4, along with the piping for the distribution of coal plant flue gas to the 8 sections and collection of the CO_2-rich anode exhaust gas from the 8 sections. The remainder of the balance-of-plant equipment such as the compressors and chillers has been centralized and located in an area to the side of the 8 ECM enclosures and opposite to the FGD unit (not-shown). Including access ways and the centralized equipment, the CEPACS system sized for 90% CO_2 capture from a 550 MW PC plant is estimated to require ~12 acres.

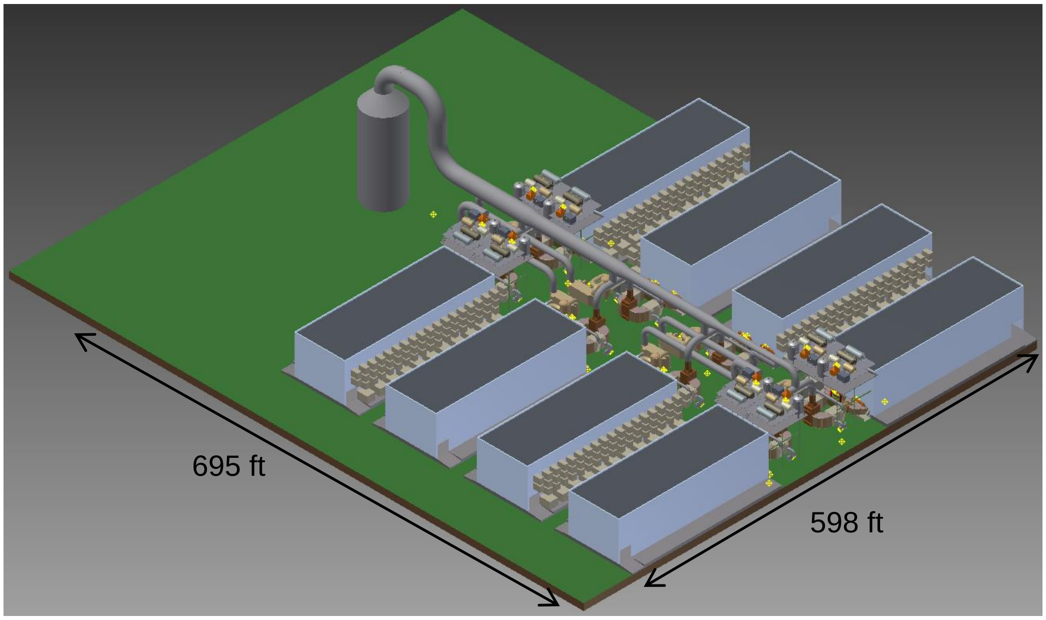


Figure 8-4. General Arrangement of ECM enclosures, BOP equipment and piping

8.2 System and Economic Analyses

Based on system simulation results, the CEPACS system performance was estimated. The ECM-based CEPACS system applied to the 550 MW PC plant simultaneously generates additional (net AC) power (after compensating for the auxiliary power requirements of CO_2 capture and compression) while capturing >90% of CO_2 from the flue gas. A list of the cases studied (evaluated/compared) is presented in Table 8-1.

Table 8-1. Case Summary

Case ID	DOE/NETL-2010/1397 Case ID [2]	Unit Cycle	MW gross	MW net	Boiler Technology	Sulfur Removal/ Recovery	CO_2 Separation
1	9	PC	583	550	Subcritical PC	Wet Flue Gas Desulfurization (FGD)	--
2	10	PC + Amine	673	550	Subcritical PC	Wet FGD	Amine Absorber
3	--	PC + ECM	1016	901	Subcritical PC	Wet FGD	ECM-based CEPACS

Assumptions made regarding the key operational and performance parameters for major subsystems or system components are listed in Table 8-2.

Table 8-2. Key System Assumptions

Case ID	**1**	**2**	**3**
Description	**Subcritical PC**	**Subcritical PC w/ CO_2 Capture (Amine)**	**Subcritical PC w/ CO_2 Capture (CEPACS)**
DOE/NETL-2010/1397 Case ID	9	10	--
Steam Cycle, MPa/°C/°C (psig/°F/°F)	16.5/566/566 (2400/1050/1050)		
Coal	Illinois No. 6		
Condenser pressure, mm Hg (in Hg)	50.8 (2)		
Boiler Efficiency, %	88		
Cooling water to condensers, °C (ºF)	16 (60)		
Cooling water from condensers, °C (ºF)	27 (80)		
Stack temperature, °C (°F)	57 (135)	32 (89)	90 (194)
SO_2 Control	Wet Limestone, Forced Oxidation		
FGD Efficiency, % (A)	98	98 (B)	Bulk: 98 Polishing: 99 (C)
NOx Control	LNB w/OFA and SCR	LNB w/OFA and SCR	LNB w/OFA, SCR, and ECM
SCR Efficiency, % (A)	86	86	86
ECM NOx Removal Efficiency, %	--	--	70
Particulate Control	Fabric Filter	Fabric Filter	Fabric Filters
Fabric Filter efficiency, % (A)	99.8	99.8	99.9
Ash Distribution, Fly/Bottom	80% / 20%	80% / 20%	80% / 20%
Mercury Control	Co-benefit Capture	Co-benefit Capture	Co-benefit Capture
Mercury removal efficiency, % (A)	90	90	90
CO_2 Control	N/A	Econamine	ECM-based CEPACS
Overall CO_2 Capture (D)	N/A	90.2%	93.0%
CO_2 Sequestration	N/A	Off-site Saline Formation	Off-site Saline Formation

A. Removal efficiencies are based on the FG content
B. An SO_2 polishing step is included to meet more stringent SO_x content limits in the FG (< 10 ppmv) to reduce formation of amine HSS (Heat Stable Salts) during the CO_2 absorption process. SO_2 exiting the post-FGD polishing step is absorbed in the CO_2 capture process making stack emissions negligible.
C. Based on remaining SO_2 in FG after bulk polishing
D. Defined as: 1-[(Stack Gas Carbon-Air Carbon)/(Total Carbon In-Air Carbon)] and expressed as percentage.

The detailed performance of the CEPACS plant, designed for (or retrofitted to) a 550 MW PC plant (from Case 9 of the referenced DOE/NETL report [26]), is summarized in Table 8-3, which includes auxiliary power requirements (including CO_2 compression).

Table 8-3. PC Plant with CEPACS CO_2 Capture System: Performance Summary

Plant Performance Summary - Case 3 - PC + ECM-based CEPACS	
CEPACS System (ECM Stack) Power, Gross AC kWe[1]	433,464
CEPACS System Auxiliary Load Summary, kWe	
Flue Gas Blower	8,126
Supplemental Air Blower	5,964
Polishing FGD	4,500
Anode Exit Compressor - Stage 1	27,369
Anode Exit Compressor - Stage 2	27,860
Chillers	397
CO_2 Liquid Pump	5,565
Water Pumps	23
Miscellaneous Balance of Plant[2]	1,500
Transformer Losses	1,355
TOTAL CEPACS AUXILIARIES, kWe	**82,659**
CEPACS NET POWER, kWe	**350,805**
CEPACS Net Plant Efficiency (HHV Natural Gas)	42.4%
CEPACS Net Plant Heat Rate, kJ/kWh	8,482
PULVERIZED COAL PLANT NET POWER, AC kWe[3]	**550,020**
TOTAL (PC + CEPACS) NET POWER w/ CO_2 capture, kWe	**900,825**
Net Plant Efficiency w/ CO_2 Capture[4]	38.8%
Net Plant Heat Rate w/ CO_2 Capture, kJ/kWh[4]	9,279
CONSUMABLES	
As-Received Coal Feed, kg/hr	198,391
Coal Thermal Input, kWt[5]	1,495,379
Natural Gas Feed, kg/hr	56,591
Natural Gas Thermal Input, kWt[6]	826,564

1. ECM Stack power includes DC-AC losses, assuming 96% conversion efficiency
2. Includes plant control systems, lighting, HVAC, and misc. low voltage loads
3. Net power from Case 9 of referenced DOE report
4. Basis: HHV Coal + HHV Natural Gas
5. HHV of As-Received Illinois No. 6 coal is 27,135 kJ/kg
6. HHV of Natural Gas is 52,581 kJ/kg

In the process of capturing ≥90% of the CO_2 from the PC plant flue gas, the ECM-based CEPACS system for the 550 MW (net AC) PC plant simultaneously generates 351 MW additional (net AC) power after compensating for the auxiliary power requirements of CO_2 capture and compression. The net electrical efficiency of the CEPACS-equipped PC plant (with CO_2 capture) was estimated to be 38.8% (based on higher heating values of coal and natural gas fuels used by PC plant and CEPACS system, respectively).

The cost estimate for the CO_2 capture system was generated utilizing the guidelines set forth in the DOE - NETL document [3]. The approach taken, and the assumptions used, in the economic feasibility study have been summarized in Section 2 earlier. Cost estimates for all major process equipment and systems were generated from a combination of vendor contacts and AECOM's historical cost databases. The capital cost for Case 3 (PC + CEPACS-based CO_2 capture) includes two components: the CEPACS system cost and the PC plant cost. The PC plant cost estimate was taken directly from the Case 9 estimate as reported in the DOE Baseline Bituminous report [26]. The capital cost estimate for the CEPACS plant was developed by AECOM in 2013/14 dollars. The TOC (total overnight cost) was then de-escalated to 2007 dollars. Table 8-4 shows the complete Case 3 plant capital cost summary (PC + CEPACS) organized by detailed cost accounts along with owner's costs, TOC, and TASC (total as-spent cost). Cost savings of approximately $20 million were estimated based on redundancy of equipment and utilities when combining the cost estimate for the CEPACS Plant (layout described in Section 8.1) with Case 9 of the DOE report. These estimated savings are primarily attributed to synergies that can be realized by centrally locating redundant equipment systems such as Limestone Feeders/Conveyors/Day Bins, Ball Mills, Dewatering equipment, electrical distribution and plant utilities. These cost savings were deducted from the combined cost estimate. The values in Table 8-4 are presented in 2007 dollars. The estimated TOC of the subcritical PC power plant with the CEPACS system for CO_2 capture is $2,297/kW.

Table 8-4. Case 3 (PC + CEPACS) Capital Cost Estimate, 2007 USD

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	$3,508,000	$0	$1,602,000	$0		$5,110,000	$480,340	$0	$838,551	$6,428,891	$7
1.2	Coal Stackout & Reclaim	$4,533,000	$0	$1,027,000	$0		$5,560,000	$522,640	$0	$912,396	$6,995,036	$8
1.3	Coal Conveyors	$4,215,000	$0	$1,016,000	$0		$5,231,000	$491,714	$0	$858,407	$6,581,121	$7
1.4	Other Coal Handling	$1,103,000	$0	$235,000	$0		$1,338,000	$125,772	$0	$219,566	$1,683,338	$2
1.5	Sorbent Receive & Unload	$140,000	$0	$42,000	$0		$182,000	$17,108	$0	$29,866	$228,974	$0
1.6	Sorbent Stackout & Reclaim	$1,469,000	$0	$269,327	$0		$1,738,327	$163,403	$0	$440,609	$2,342,339	$3
1.7	Sorbent Conveyors	$510,000	$110,185	$125,296	$0		$745,481	$70,075	$0	$194,294	$1,009,851	$1
1.8	Other Sorbent Handling	$239,000	$56,207	$125,609	$0		$420,816	$39,557	$0	$141,290	$601,663	$1
1.9	Coal & Sorbent Hnd.Foundations	$0	$2,058,000	$2,596,386	$0		$4,654,386	$437,512	$0	$1,598,826	$6,690,724	$7
	SUBTOTAL 1.	**$15,717,000**	**$2,224,392**	**$7,038,619**	**$0**	**$0**	**$24,980,011**	**$2,348,121**	**$0**	**$5,233,805**	**$32,561,937**	**$36**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$2,014,000	$0	$393,000	$0		$2,407,000	$226,258	$0	$394,989	$3,028,247	$3
2.2	Coal Conveyor to Storage	$5,158,000	$0	$1,126,000	$0		$6,284,000	$590,696	$0	$1,031,204	$7,905,900	$9
2.3	Coal Injection System	$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
2.5	Sorbent Prep Equipment	$2,657,000	$114,354	$551,791	$0		$3,323,144	$312,376	$0	$791,618	$4,427,138	$5
2.6	Sorbent Storage & Feed	$438,929	$6,198	$116,768	$12,953		$574,848	$54,036	$16,237	$161,243	$806,363	$1
2.7	Sorbent Injection System	$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
2.9	Coal & Sorbent Feed Foundation	$0	$299,000	$251,064	$0		$550,064	$51,706	$0	$150,644	$752,414	$1
	SUBTOTAL 2.	**$10,267,929**	**$419,552**	**$2,438,622**	**$12,953**	**$0**	**$13,139,056**	**$1,235,071**	**$16,237**	**$2,529,698**	**$16,920,063**	**$19**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	Feedwater System	$13,029,496	$15,940	$4,500,719	$91,339		$17,637,494	$1,657,924	$25,365	$3,675,631	$22,996,415	$26
3.2	Water Makeup & Pretreating	$3,814,277	$312,500	$1,300,234	$104,279		$5,531,290	$519,941	$35,072	$1,494,179	$7,580,482	$8
3.3	Other Feedwater Subsystems	$5,318,000	$0	$2,248,000	$0		$7,566,000	$711,204	$0	$1,241,699	$9,518,903	$11
3.4	Service Water Systems	$931,000	$0	$507,000	$0		$1,438,000	$135,172	$0	$314,634	$1,887,806	$2
3.5	Other Boiler Plant Systems	$6,634,980	$458,653	$6,398,926	$258,174		$13,750,733	$1,292,569	$65,637	$2,266,418	$17,375,356	$19
3.6	FO Supply Sys & Nat Gas	$4,613,089	$59,789	$635,459	$370,322		$5,678,658	$533,794	$255,133	$970,368	$7,437,953	$8
3.7	Waste Treatment Equipment	$3,221,000	$0	$1,836,000	$0		$5,057,000	$475,358	$0	$1,106,598	$6,638,956	$7
3.8	Misc. Equip.(cranes, AirComp., Comm.)	$1,965,000	$0	$599,994	$0		$2,564,994	$241,109	$0	$775,611	$3,581,715	$4
	SUBTOTAL 3.	**$39,526,842**	**$846,882**	**$18,026,332**	**$824,114**	**$0**	**$59,224,170**	**$5,567,072**	**$381,207**	**$11,845,138**	**$77,017,587**	**$85**
4	PC BOILER & ACCESSORIES											
4.1	PC Boiler & Accessories	$134,824,000	$0	$86,704,000	$0		$221,528,000	$20,823,632	$0	$24,235,163	$266,586,795	$296
4.2	SCR (w/4.1)	$0	$0	$0	$0		$0	$0	$0	$0	$0	$0
4.3	Open	$0	$0	$0	$0		$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$130,609	$0	$60,824	$71,402		$262,834	$24,706	$13,142	$45,102	$345,784	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$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
4.8	Major Component Rigging		w/4.1	w/4.1	$0		$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations		w/14.1	w/14.1	$0		$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$134,954,609**	**$0**	**$86,764,824**	**$71,402**	**$0**	**$221,790,834**	**$20,848,338**	**$13,142**	**$24,280,265**	**$266,932,579**	**$296**

Table 8-4 Case 3 (PC + CEPACS) Capital Cost Estimate, 2007 USD (continued)

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
5	FLUE GAS CLEANUP											
5.1	Absorber Vessels & Accessories	$58,362,000	$0	$12,564,000	$0		$70,926,000	$6,667,044	$0	$7,759,304	$85,352,348	$95
5.2	Other FGD	$16,881,481	$192,648	$4,251,594	$939,828		$22,265,551	$2,092,962	$788,428	$3,416,655	$28,563,595	$32
5.3	Bag House & Accessories	$18,374,600	$309,476	$11,714,186	$1,464,088		$31,862,350	$2,995,061	$245,118	$3,790,667	$38,893,196	$43
5.4	Other Particulate Removal Materials	$2,210,878	$131,245	$1,510,937	$372,057		$4,225,118	$397,161	$95,756	$581,348	$5,299,383	$6
5.5	Gypsum Dewatering System	$2,532,000	$0	$429,942	$0		$2,961,942	$278,423	$0	$611,984	$3,852,348	$4
5.6	Mercury Removal System	$0	$0	$0	$0		$0	$0	$0	$0	$0	$0
5.7	Polishing FGD	$59,642,529	$0	$4,385,096	$5,147,721		$69,175,346	$6,502,483	$3,458,767	$11,870,489	$91,007,086	$101
5.9	Open	$0	$0	$0	$0		$0	$0	$0	$0	$0	$0
	SUBTOTAL 5.	**$158,003,488**	**$633,369**	**$34,855,756**	**$7,923,694**	**$0**	**$201,416,308**	**$18,933,133**	**$4,588,068**	**$28,030,448**	**$252,967,957**	**$281**
5B	CO_2 REMOVAL & COMPRESSION											
5B.1	CO_2 Removal System	$19,323,463	$87,600	$303,907	$356,760		$20,071,730	$1,886,743	$1,003,587	$3,444,309	$26,406,368	$29
5B.2	CO_2 Compression & Drying	$52,602,856	$770,273	$2,121,332	$2,490,260		$57,984,722	$5,450,564	$2,899,236	$9,950,178	$76,284,700	$85
	SUBTOTAL 5B.	**$71,926,320**	**$857,874**	**$2,425,239**	**$2,847,020**	**$0**	**$78,056,452**	**$7,337,306**	**$3,902,823**	**$13,394,487**	**$102,691,068**	**$114**
5C	Electrochemical Module											
5C.1	Electrochemical Module	$202,446,566	$16,907,188	$3,570,951	$4,191,986		$227,116,691	$21,348,969	$11,355,835	$38,973,224	$298,794,719	$332
5C.2	Electrochemical Module Civil	$0	$7,688,113	$4,341,611	$5,096,674		$17,126,397	$1,609,881	$856,320	$2,938,890	$22,531,489	$25
5C.3	Electrochemical Module Piping	$0	$0	$0	$0		$0	$0	$0	$0	$0	$0
	SUBTOTAL 5C	**$202,446,566**	**$24,595,301**	**$7,912,562**	**$9,288,660**	**$0**	**$244,243,089**	**$22,958,850**	**$12,212,154**	**$41,912,114**	**$321,326,208**	**$357**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A		N/A			$0	$0	$0	$0	$0	$0
6.2	Open						$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping						$0	$0	$0	$0	$0	$0
6.4	Combustion Turbine Foundations						$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
7.2	HRSG System	$0	$0	$0	$0		$0	$0	$0	$0	$0	$0
7.3	Ductwork	$9,097,000	$0	$5,845,000	$0		$14,942,000	$1,404,548	$0	$2,451,982	$18,798,530	$21
7.4	Stack	$9,145,000	$0	$5,351,000	$0		$14,496,000	$1,362,624	$0	$1,585,862	$17,444,486	$19
7.9	Duct & Stack Foundations	$0	$1,049,000	$1,192,000	$0		$2,241,000	$210,654	$0	$489,938	$2,941,592	$3
	SUBTOTAL 7.	**$18,242,000**	**$1,049,000**	**$12,388,000**	**$0**	**$0**	**$31,679,000**	**$2,977,826**	**$0**	**$4,527,783**	**$39,184,609**	**$43**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$49,912,000	$0	$6,242,000	$0	$0	$56,154,000	$5,278,476	$0	$6,142,796	$67,575,272	$75
8.2	Turbine Plant Auxiliaries	$348,000	$0	$746,000	$0	$0	$1,094,000	$102,836	$0	$119,325	$1,316,161	$1
8.3	Condenser & Auxiliaries	$8,435,976	$3,699	$2,305,533	$12,364	$0	$10,757,572	$1,011,212	$60,579	$1,252,567	$13,081,929	$15
8.4	Steam Piping	$16,005,000	$0	$7,891,000	$0	$0	$23,896,000	$2,246,224	$0	$3,920,896	$30,063,120	$33
8.9	TG Foundations	$0	$1,092,000	$1,726,000	$0	$0	$2,818,000	$264,892	$0	$616,398	$3,699,290	$4
	SUBTOTAL 8.	**$74,700,976**	**$1,095,699**	**$18,910,533**	**$12,364**	**$0**	**$94,719,572**	**$8,903,640**	**$60,579**	**$12,051,982**	**$115,735,772**	**$128**

Table 8-4 Case 3 (PC + CEPACS) Capital Cost Estimate, 2007 USD (continued)

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
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$13,033,000	$0	$4,057,908	$0	$0	$17,090,908	$1,606,545	$0	$1,403,493	$20,100,946	$22
9.2	Circulating Water Pumps	$2,783,000	$0	$175,221	$0	$0	$2,958,221	$278,073	$0	$236,296	$3,472,590	$4
9.3	Circ.Water System Auxiliaries	$531,000	$0	$71,000	$0	$0	$602,000	$56,588	$0	$65,796	$724,384	$1
9.4	Circ.Water Piping	$0	$4,210,000	$4,080,000	$0	$0	$8,290,000	$779,260	$0	$1,360,357	$10,429,617	$12
9.5	Make-up Water System	$462,000	$0	$618,000	$0	$0	$1,080,000	$101,520	$0	$177,126	$1,358,646	$2
9.6	Component Cooling Water Sys	$421,000	$0	$335,000	$0	$0	$756,000	$71,064	$0	$124,028	$951,092	$1
9.9	Circ.Water System Foundations & Structures	$0	$2,500,000	$3,972,000	$0	$0	$6,472,000	$608,368	$0	$1,416,074	$8,496,442	$9
	SUBTOTAL 9.	**$17,230,000**	**$6,710,000**	**$13,309,129**	**$0**	**$0**	**$37,249,129**	**$3,501,418**	**$0**	**$4,783,170**	**$45,533,717**	**$51**
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	$612,000	$0	$1,885,000	$0	$0	$2,497,000	$234,718	$0	$273,172	$3,004,890	$3
10.7	Ash Transport & Feed Equipment	$3,961,000	$0	$4,058,000	$0	$0	$8,019,000	$753,786	$0	$877,279	$9,650,065	$11
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$0	$171,000	$0	$0	$171,000	$16,074	$0	$37,415	$224,489	$0
	SUBTOTAL 10.	**$4,573,000**	**$0**	**$6,114,000**	**$0**	**$0**	**$10,687,000**	**$1,004,578**	**$0**	**$1,187,865**	**$12,879,443**	**$14**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$1,602,000	$0	$260,000	$0	$0	$1,862,000	$175,028	$0	$203,703	$2,240,731	$2
11.2	Station Service Equipment	$2,904,000	$0	$954,000	$0	$0	$3,858,000	$362,652	$0	$422,065	$4,642,717	$5
11.3	Switchgear & Motor Control	$3,339,000	$6,585,692	$1,054,405	$572,171	$0	$11,551,268	$1,085,819	$382,263	$1,739,470	$14,758,821	$16
11.4	Conduit & Cable Tray	$0	$3,039,462	$7,354,064	$136,249	$0	$10,529,775	$989,799	$59,939	$1,736,842	$13,316,355	$15
11.5	Wire & Cable	$0	$3,950,000	$7,625,000	$0	$0	$11,575,000	$1,088,050	$0	$1,899,514	$14,562,564	$16
11.6	Protective Equipment	$270,000	$0	$918,000	$0	$0	$1,188,000	$111,672	$0	$129,569	$1,429,241	$2
11.7	Standby Equipment	$1,279,000	$0	$29,000	$0	$0	$1,308,000	$122,952	$0	$143,189	$1,574,141	$2
11.8	Main Power Transformers	$8,414,000	$11,184,595	$1,108,992	$1,099,947	$0	$21,807,534	$2,049,908	$661,077	$3,208,348	$27,726,866	$31
11.9	Electrical Foundations	$0	$312,000	$765,000	$0	$0	$1,077,000	$101,238	$0	$235,648	$1,413,886	$2
	SUBTOTAL 11.	**$17,808,000**	**$25,071,749**	**$20,068,461**	**$1,808,367**	**$0**	**$64,756,577**	**$6,087,118**	**$1,103,279**	**$9,718,348**	**$81,665,322**	**$91**
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	$1,143,359	$274,878	$322,683	$0	$1,740,920	$163,646	$87,046	$298,742	$2,290,354	$3
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	$446,000	$0	$267,000	$0	$0	$713,000	$67,022	$0	$117,003	$897,025	$1
12.7	Distributed Control System Equipment	$4,504,000	$2,744,000	$787,000	$0	$0	$8,035,000	$755,290	$137,200	$1,049,706	$9,977,196	$11
12.8	Instrument Wiring & Tubing	$2,442,000	$0	$4,844,000	$0	$0	$7,286,000	$684,884	$0	$1,195,633	$9,166,517	$10
12.9	Other I & C Equipment	$1,273,000	$0	$2,888,000	$0	$0	$4,161,000	$391,134	$0	$455,213	$5,007,347	$6
	SUBTOTAL 12.	**$8,665,000**	**$3,887,359**	**$9,060,878**	**$322,683**	**$0**	**$21,935,920**	**$2,061,976**	**$224,246**	**$3,116,297**	**$27,338,439**	**$30**

Table 8-4 Case 3 (PC + CEPACS) Capital Cost Estimate, 2007 USD (continued)

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
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$3,512,262	$2,141,035	$1,339,476	$0	$6,992,774	$657,321	$297,139	$1,250,051	$9,197,285	$10
13.2	Site Improvements	$0	$3,330,674	$5,066,625	$3,528,342	$0	$11,925,641	$1,121,010	$410,232	$2,221,717	$15,678,601	$17
13.3	Site Facilities	$2,974,000	$0	$2,933,000	$0	$0	$5,907,000	$555,258	$0	$1,292,097	$7,754,355	$9
	SUBTOTAL 13.	**$2,974,000**	**$6,842,936**	**$10,140,660**	**$4,867,819**	**$0**	**$24,825,415**	**$2,333,589**	**$707,371**	**$4,763,866**	**$32,630,241**	**$36**
14	BUILDINGS & STRUCTURES											
14.1	Boiler Building	$0	$8,519,000	$7,491,000	$0	$0	$16,010,000	$1,504,940	$0	$3,502,988	$21,017,928	$23
14.2	Turbine Building	$0	$12,310,000	$11,473,000	$0	$0	$23,783,000	$2,235,602	$0	$5,203,720	$31,222,322	$35
14.3	Administration Building	$0	$587,000	$621,000	$0	$0	$1,208,000	$113,552	$0	$264,310	$1,585,862	$2
14.4	Circulation Water Pumphouse	$0	$168,000	$134,000	$0	$0	$302,000	$28,388	$0	$66,078	$396,466	$0
14.5	Water Treatment Buildings	$0	$603,000	$549,000	$0	$0	$1,152,000	$108,288	$0	$252,058	$1,512,346	$2
14.6	Machine Shop	$0	$393,000	$264,000	$0	$0	$657,000	$61,758	$0	$143,752	$862,510	$1
14.7	Warehouse	$0	$266,000	$267,000	$0	$0	$533,000	$50,102	$0	$116,620	$699,722	$1
14.8	Other Buildings & Structures	$0	$5,442,102	$2,802,704	$3,072,956	$0	$11,317,762	$1,063,870	$545,788	$1,961,102	$14,888,522	$17
14.9	Waste Treating Building & Str.	$0	$416,000	$1,263,000	$0	$0	$1,679,000	$157,826	$0	$367,365	$2,204,191	$2
	SUBTOTAL 14.	**$0**	**$28,704,102**	**$24,864,704**	**$3,072,956**	**$0**	**$56,641,762**	**$5,324,326**	**$545,788**	**$11,877,994**	**$74,389,869**	**$83**
15	Piping and Ductwork											
15.1	Piping and Ductwork	$0	$44,713,922	$47,565,004	$55,837,179	$0	$148,116,105	$13,922,914	$7,405,805	$25,416,724	$194,861,548	$216
	SUBTOTAL 15.	**$0**	**$44,713,922**	**$47,565,004**	**$55,837,179**	**$0**	**$148,116,105**	**$13,922,914**	**$7,405,805**	**$25,416,724**	**$194,861,548**	**$216**
	TOTAL COST	**$777,035,729**	**$147,652,136**	**$321,883,323**	**$86,889,211**	**$0**	**$1,333,460,399**	**$125,345,278**	**$31,160,699**	**$204,669,984**	**$1,694,636,359**	**$1,881**
	Owner's Costs											
	Preproduction Costs											
	6 Months All Labor										$7,985,608	$9
	1 Month Maintenance Materials										$1,629,326	$2
	1 Month Non-fuel Consumables										$963,840	$1
	1 Month Waste Disposal										$251,000	$0
	Operator Training/Plant Personel Project Support										$73,382	$0
	25% of 1 Months Fuel Cost at 100% CF										$4,677,830	$5
	2% of TPC										$33,892,727	$38
	Total										**$49,473,713**	**$55**
	Inventory Capital											
	60 day supply of fuel and consumables at 100% CF										$13,839,680	$15
	0.5% of TPC (spare parts)										$8,473,182	$9
	Total										**$22,312,862**	**$25**
	Initial Cost for Catalyst and Chemicals										$1,749,346	$2
	Land										$942,900	$1
	Other Owner's Costs										$254,195,454	$282
	Financing Costs										$45,755,182	$51
	Total Overnight Costs (TOC)										**$2,069,065,816**	**$2,297**
	TASC Multiplier										1.140	
	Total As-Spent Cost (TASC)										**$2,358,735,030**	**$2,618**

The cost of electricity was estimated based on the NETL cost estimation methodology guidelines [3]. Tables 8-5 and 8-6 show the Fixed (OC_{FIX}) and Variable (OC_{VAR}) operating costs, respectively, for the CEPACS system. The estimates were developed by AECOM in a combination of 2012 and 2013/14 dollars and de-escalated to 2007 dollars. Dollar year basis for estimates are listed for each account (e.g.: fixed operating costs in 2012 USD, variable operating costs in 2014 USD, etc.). All variable costs were estimated as a function of the capacity factor.

Table 8-5. Fixed Operating Costs for CEPACS System

Parameter	Annual Cost	Notes
Annual Operating Labor Cost/ Maintenance Labor Cost	$ 2,249,000	1 operator + 2 assistants for each shift. 4 maint. staff on day shift. Rate = 65$/h.
Property Taxes and Insurance	$ 363,175	$995 / calendar day
Total CEPACS Fixed Annual Operating Costs (OC_{FIX}) 2012 USD	**$ 2,612,175**	
Total CEPACS Plant OC_{FIX}, 2007 USD	$ 2,253,285	
Total PC Plant (Case 1) OC_{FIX}, 2007 USD	$ 32,056,744	
Total Case 3 (PC + CEPACS) OC_{FIX}, 2007 USD	**$ 34,310,029**	

Table 8-6. Variable Operating Costs for CEPACS System

Parameter	Initial Fill Cost	Operating Hours	Replacement Cost	Annual Cost
Maintenance Material Cost				$ 7,751,604
Consumables				
Consumables, Limestone				$ 114,750
Process Water (generation credit)				$ (101,493)
Others - Catalysts				
Sorbent - Desulfurizers	$ 2,394,855	9464 hrs	$ 1,197,427	$ 902,186
ECM Stack Replacement	$ 227,520,868	10 years	$ 352,657,345	$ 11,755,244.8
Total CEPACS Variable Annual Operating Costs (OC_{VAR}) 2014 USD				**$ 20,422,292**
Total CEPACS Plant OC_{VAR}, 2007 USD				**$ 15,943,593**
Total PC Plant (Case 1) OC_{VAR}, 2007 USD				**$ 21,078,354**
Total Case 3 (PC + CEPACS) OC_{VAR}, 2007 USD				**$ 37,021,947**

Table 8-7 shows the cost of electricity and cost of carbon dioxide captured and avoided for Case 3 (PC + CEPACS-based CO_2 capture) as compared to Case 1 (PC w/o CO_2 capture) and Case 2 (PC w/ Amine-based CO_2 capture). The estimated COE for Case 3 is 80.9 mills/kWh and the cost of CO_2 captured is \$33.63/tonne. The cost of CO_2 avoided was estimated to be \$26.67/tonne for Case 3.

Table 8-7. Case 3 (PC + CEPACS) COE and Cost of CO_2 Captured, Compared to Cases 1 and 2

	Case 1 PC	Case 2 PC w/ Amine CO2 cap.	Case 3 PC + CEPACS
TOC	\$ 1,098,124,000	\$ 1,985,432,000	\$ 2,069,065,816
Capital Charge Factor (CCF)	11.65%	12.40%	12.40%
OC_{FIX}	\$ 32,056,744	\$ 53,460,210	\$ 34,310,029
OC_{VAR}	\$ 21,078,354	\$ 37,495,580	\$ 37,021,947
Capacity Factor (CF)	85%	85%	85%
MW,net	550.020	549.960	900.825
Coal Usage, tons/day	5,248	7,380	5,248
Gas Usage, MMBTU/day	--	--	67,672
CO_2 TS&M, \$/Ton	--	\$ 3.194	\$ 3.194
CO_2 Captured, tons/year	--	4,895,935	4,727,085
Cost of Electricity			
Capital	31.2	60.1	38.3
Fixed OCs	7.8	13.1	5.1
Variable OCs	5.1	9.2	5.5
Fuel Cost	15.2	21.3	29.8
CO_2 TS&M	0	5.8	2.3
COE, mills/kWh	**59.4**	**109.5**	**80.9**
Incremental COE (Case 1 Basis)	**--**	**84.3%**	**36.2%**
Levelized COE, mills/kWh	75.3	138.8	102.6
Cost of CO_2 Capture			
Cost per Ton of CO_2 captured, (2007\$)		**\$ 41.89**	**\$ 30.52**
Cost per Ton of CO_2 avoided, (2007\$)		**\$ 60.94**	**\$ 24.20**
Cost per tonne of CO_2 captured, (2007\$)		**\$ 46.17**	**\$ 33.63**
Cost per tonne of CO_2 avoided, (2007\$)		**\$ 67.15**	**\$ 26.67**

Table 8-8 shows the cost, performance, and environmental profile summary for all cases. The results are discussed below in the following order:

- Performance (efficiency and raw water consumption)
- Cost (plant capital costs, COE, and cost per ton of CO_2 captured)
- Environmental profile (CO_2 and other emissions)

Table 8-8. Cost and Performance Summary for All Cases

	Case 1 PC w/o CO_2 capture	PC w/ Amine CO_2 Cap	PC + CEPACS
PERFORMANCE			
CO_2 Capture	0%	90%	92.8%
Gross Power Output (kWe)	582,600	672,700	1,016,064
Auxiliary Power Requirement (kWe)	32,580	122,740	115,239
Net Power Output (kWe)	550,020	549,960	900,825
Coal Flow rate (lb/hr)	437,378	614,994	437,378
Natural Gas Flow rate (lb/hr)	N/A	N/A	124,763
HHV Thermal Input (kWth)	1,495,379	2,102,643	2,321,943
Net Plant HHV Efficiency (%)	36.8%	26.2%	38.8%
Net Plant HHV Heat Rate (Btu/kWh)	9,277	13,046	8,795
Raw Water Withdrawal (gpm/MWnet)	10.7	20.4	8.1
Process Water Discharge (gpm/MWnet)	2.2	4.7	1.9
Raw Water Consumption (gpm/MWnet)	8.5	15.7	6.2
CO_2 Emissions (lb/MMBtu)	204	20	13
CO_2 Emissions (lb/MWhgross)	1,783	217	98
CO_2 Emissions (lb/MWhnet)	1,888	266	111
SO_2 Emissions (lb/MMBtu)	0.0858	0.0017	0.0006
SO_2 Emissions (lb/MWhgross)	0.7515	0.0176	0.0047
NOx Emissions (lb/MMBtu)	0.07	0.07	0.0155
NOx Emissions (lb/MWhgross)	0.613	0.747	0.1205
PM Emissions (lb/MMBtu)	0.013	0.013	0.0042
PM Emissions (lb/MWhgross)	0.114	0.139	0.0329
Hg Emissions (lb/TBtu)	1.143	1.143	0.7360
Hg Emissions (lb/MWhgross)	1.00E-05	1.22E-05	5.74E-06
COST			
Total Plant Cost (2007$/kW)	1,622	2,942	1,881
Total Overnight Cost (2007$/kW)	1,996	3,610	2,297
Bare Erected Cost (2007$/kW)	1,317	2,255	1,480
Home Office Expenses (2007$/kW)	124	213	139
Project Contingency (2007$/kW)	182	369	227
Process Contingency (2007$/kW)	0	105	35
Owner's Costs (2007$/kW)	374	667	416
Total Overnight Cost (2007$ x 1,000)	1,098,124	1,985,432	2,068,815
Total As Spent Capital (2007$/kW)	2,264	4,115	2,618
CO2 TS&M Costs (mills/kWh, 2007$)	0.0	5.8	2.3
Fuel Costs (mills/kWh, 2007$)	15.2	21.3	29.8
Variable Costs (mills/kWh, 2007$)	5.1	9.2	5.5
Fixed Costs (mills/kWh, 2007$)	7.8	13.1	5.1
Capital Costs (mills/kWh, 2007$)	31.2	60.2	38.3
COE (mills/kWh, 2007$)	**59.4**	**109.6**	**80.9**
Incremental COE, Case 1 Basis	0	84.3%	36.2%
LCOE (mills/kWh, 2007$)	75.3	139.0	102.6
Cost of CO_2 Captured ($/ton - 2007)	--	$ 41.89	$ 30.52
Cost of CO_2 Avoided ($/ton - 2007)	--	$ 60.94	$ 24.20
Cost of CO_2 Captured ($/tonne - 2007)	--	$ 46.17	$ 33.63
Cost of CO_2 Avoided ($/tonne - 2007)	--	$ 67.15	$ 26.67

Performance

The plant net efficiency (HHV basis) for all cases is presented in Figure 8-5. Plant Electrical Efficiency (HHV). The key conclusions that can be drawn are:

- The CEPACS-equipped PC plant (Case 3) has the highest net electrical efficiency at 38.8%.
- The net electrical efficiency of the CEPACS-equipped (for CO_2 capture) PC plant (Case 3) is 5.4% higher than that of the baseline PC plant without CO_2 capture (Case 1).

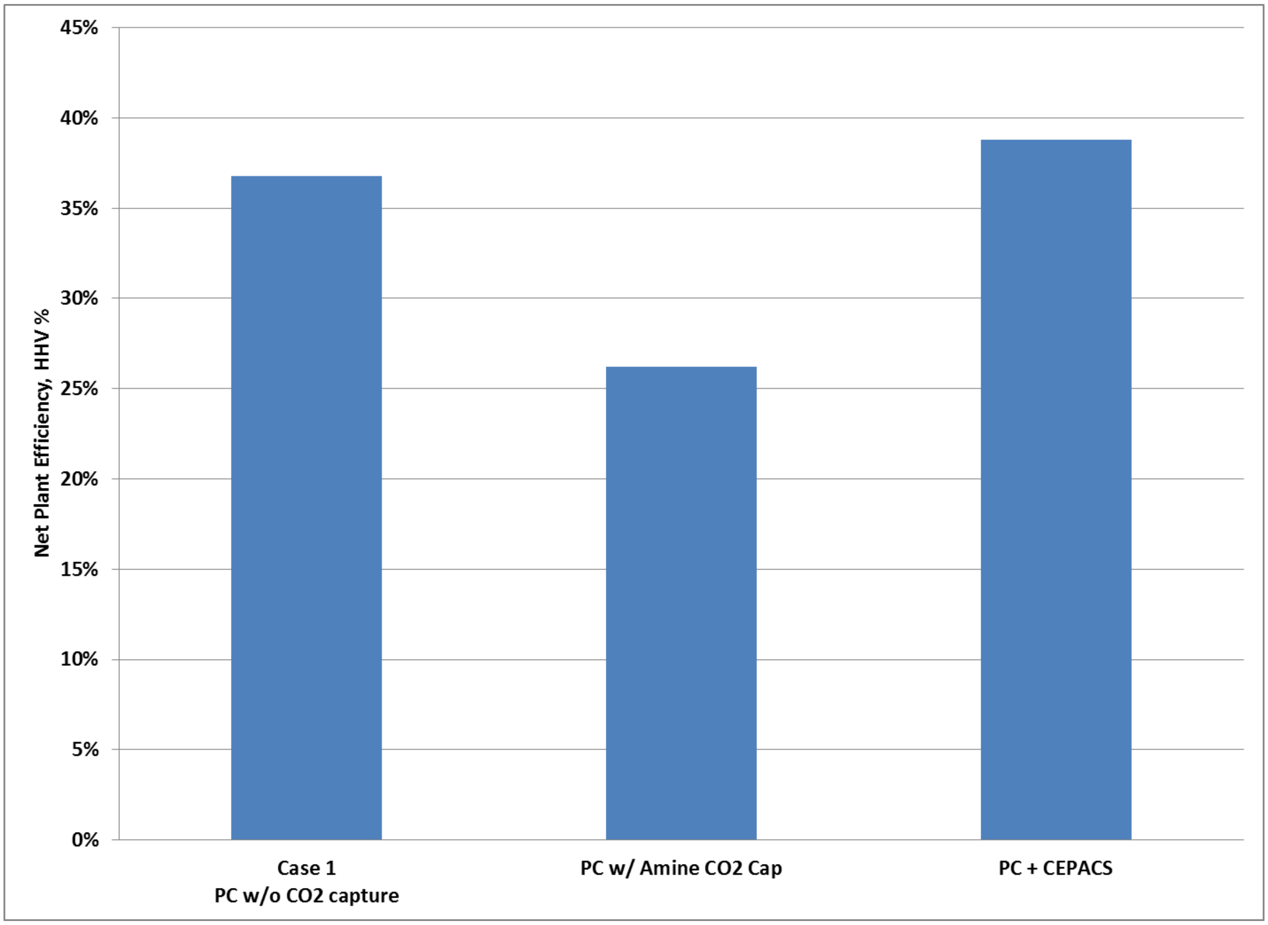


Figure 8-5. Plant Electrical Efficiency (HHV)

A comparison of water use is presented in Figure 8-6. Water Use. Three categories of water use, normalized by plant net power output, are shown: water withdrawal, discharge, and consumption. Water withdrawal is the amount of water removed from the ground or other supply source for the process. Water consumption is the amount of water that is consumed within the process (i.e.: evaporated, transpired, or incorporated into products). Water discharge is water that is treated and discharged to the environment. The primary conclusions that can be drawn are:

- The CEPACS-equipped PC plant (Case 3) has the lowest water use (withdrawal) requirements of all cases, including Case 1. Case 3 normalized water withdrawal is 24.3% less than that for Case 1 (no CO_2 capture). This is partially due to the fact that water is generated within the ECM stacks as a byproduct of the electrochemical reactions.

- The PC plant with Econamine-based CO_2 capture process (Case 2) requires 151.9% more water withdrawal than the PC plant with CEPACS-based CO_2 capture process (Case 3).

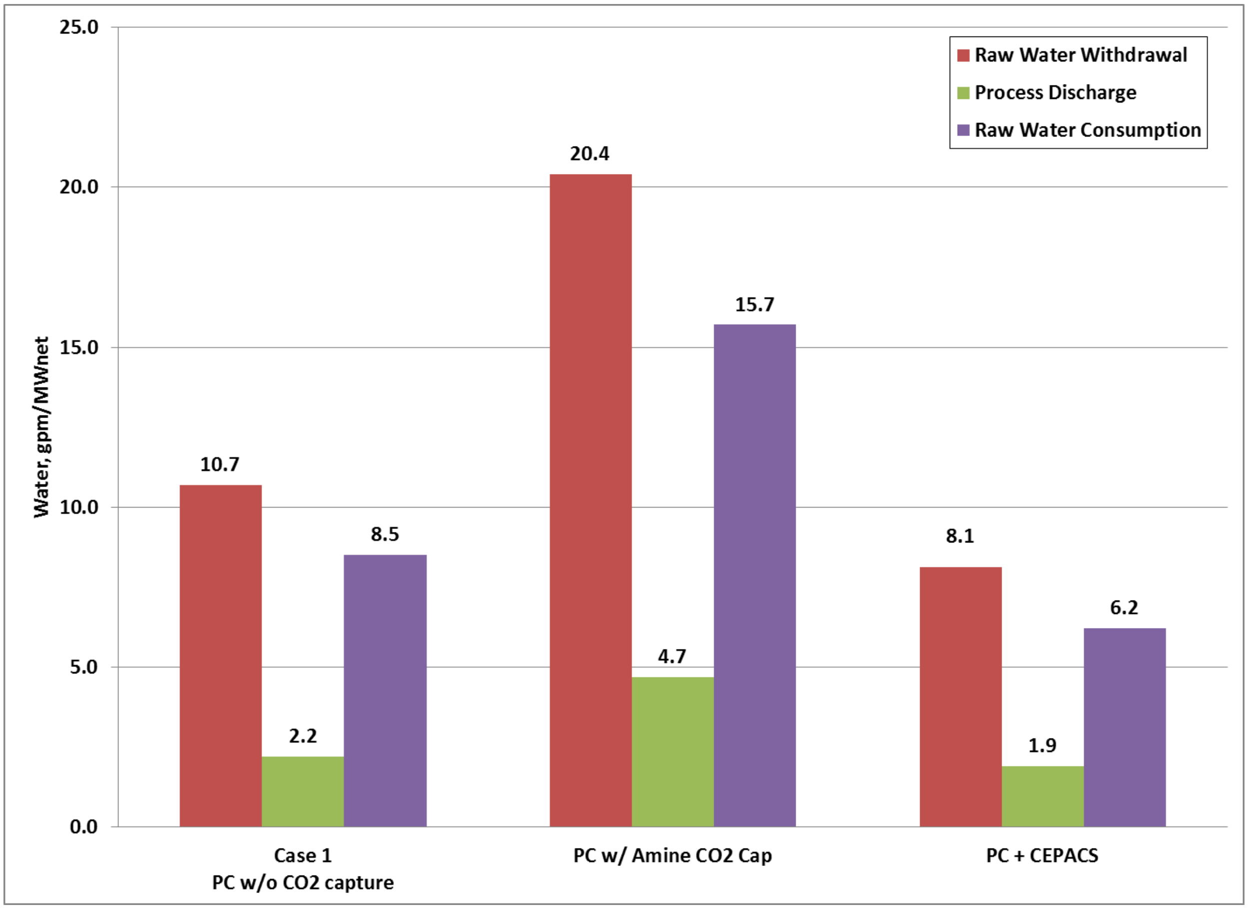


Figure 8-6. Water Use

Plant Capital Cost

The Total Overnight Cost (TOC) was estimated for each plant by adding owner's costs to the Total Plant Cost (TPC). The TPC for the CEPACS case was estimated by AECOM (formerly URS Corporation) based on a combination of vendor quotations and in-house cost databases. The TPC and TOC estimates for Cases 1 and 2 were based on the reported values in the referenced DOE/NETL Baseline Bituminous report [2].

The normalized TOC (with breakdown by component) and Total As-Spent Capital (TASC) for each case are shown in Figure 8-7. The following key conclusions can be drawn:

- Case 3 (PC + CEPACS) has the lowest TOC of the two cases with carbon capture, at $2,297/kW (2007 USD).

- The TOC for Case 3 is only 15.1% higher than the TOC for Case 1. This is because the additional costs of the CEPACS system are partially offset by the additional power generated by the system.
- The TOC of the Econamine-based system (Case 2) is 80.9% higher than the TOC for Case 1.

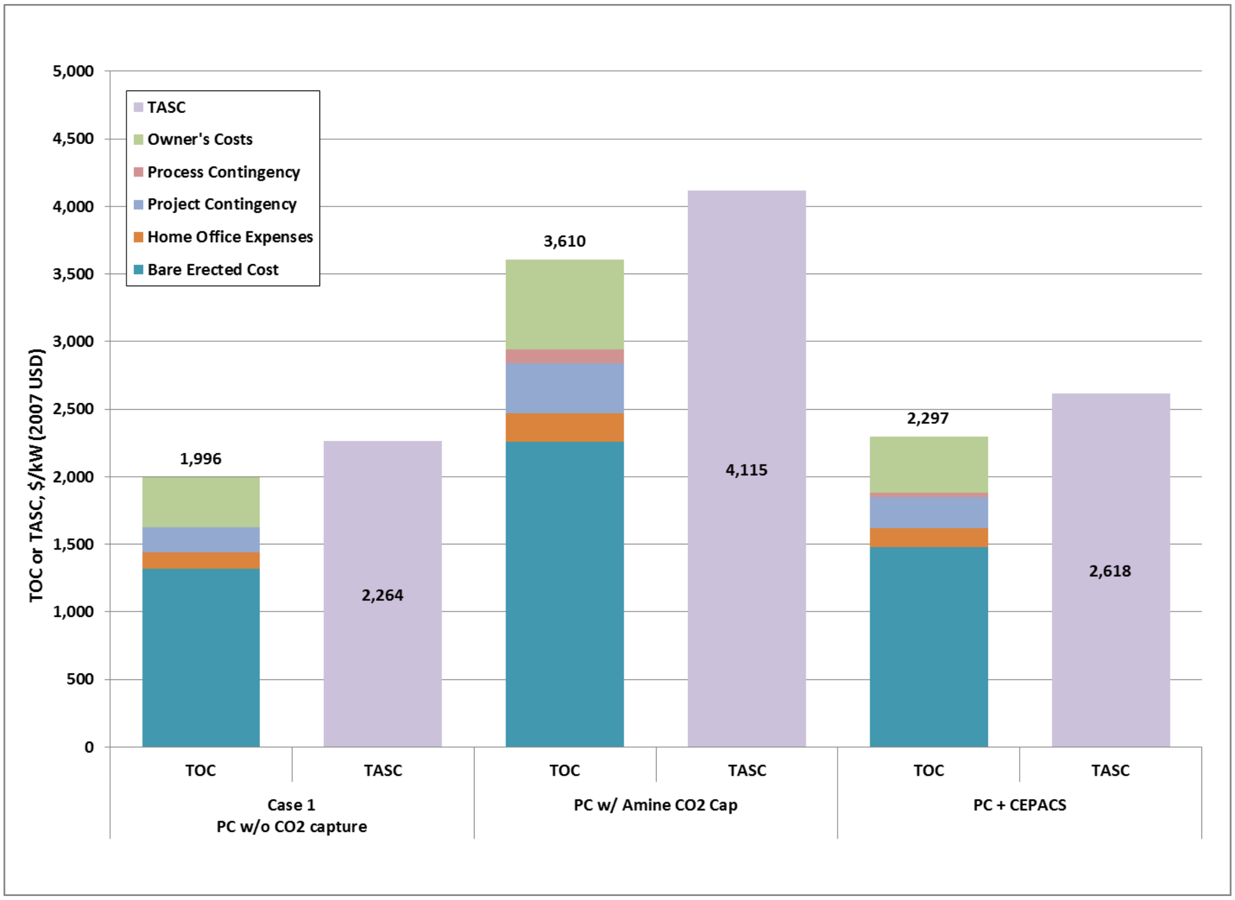


Figure 8-7. Plant Capital Costs

Cost of Electricity and CO_2 Captured

Figure 8-8 shows the Cost of Electricity (COE) with component breakdown and Incremental COE (Case 1 as basis) calculated for each case. The following key conclusion can be drawn:

- Case 3 (PC + CEPACS) offers the lowest COE of all cases with carbon capture at 80.9 mills/kWh. The incremental COE for Case 3 is 36.2%.

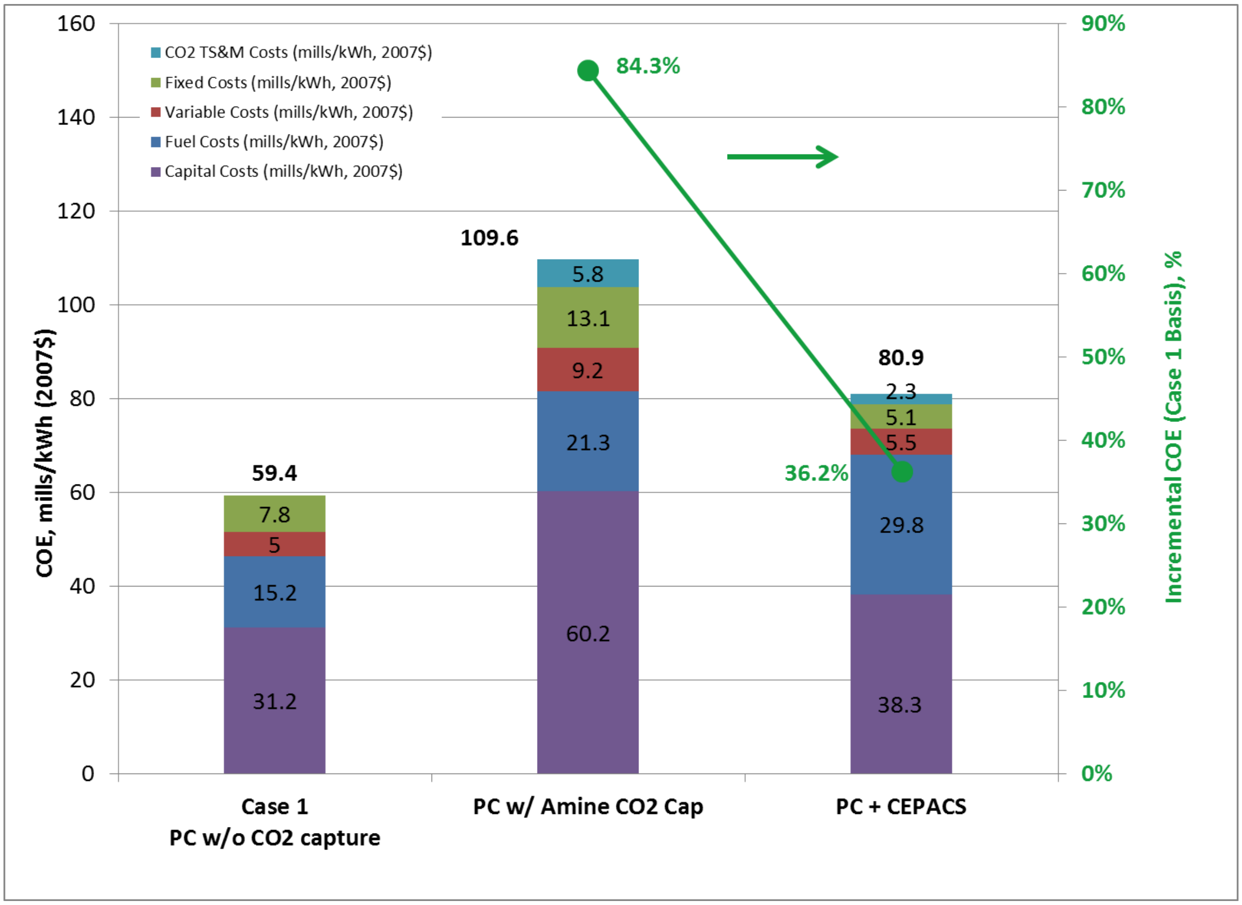


Figure 8-8. Cost of Electricity and Incremental COE

Figure 8-9 shows the cost of CO_2 captured (Case 1 as basis) calculated for the two CO_2 capture cases. The costs are shown in year 2007 US dollars and per metric ton of CO_2, for consistency with the T&EFS estimation basis. The following key conclusion can be drawn:

- The CEPACS system cost of CO_2 captured is \$33.63/tonne (in 2007 USD), which compares favorably with the DOE target of less than \$40/tonne (2011 USD).
- The cost of CO_2 captured for the CEPACS-based process (Case 3) is 27.2% lower than that for the Econamine-based process (Case 2).
- The cost of CO_2 avoided for Case 3 is 60.3% lower than that for Case 2, due to the extra power generation of the CEPACS plant which results in significantly lower CO_2 emissions (lb/MWh) than Case 2, at a fixed 90% CO_2 capture rate.

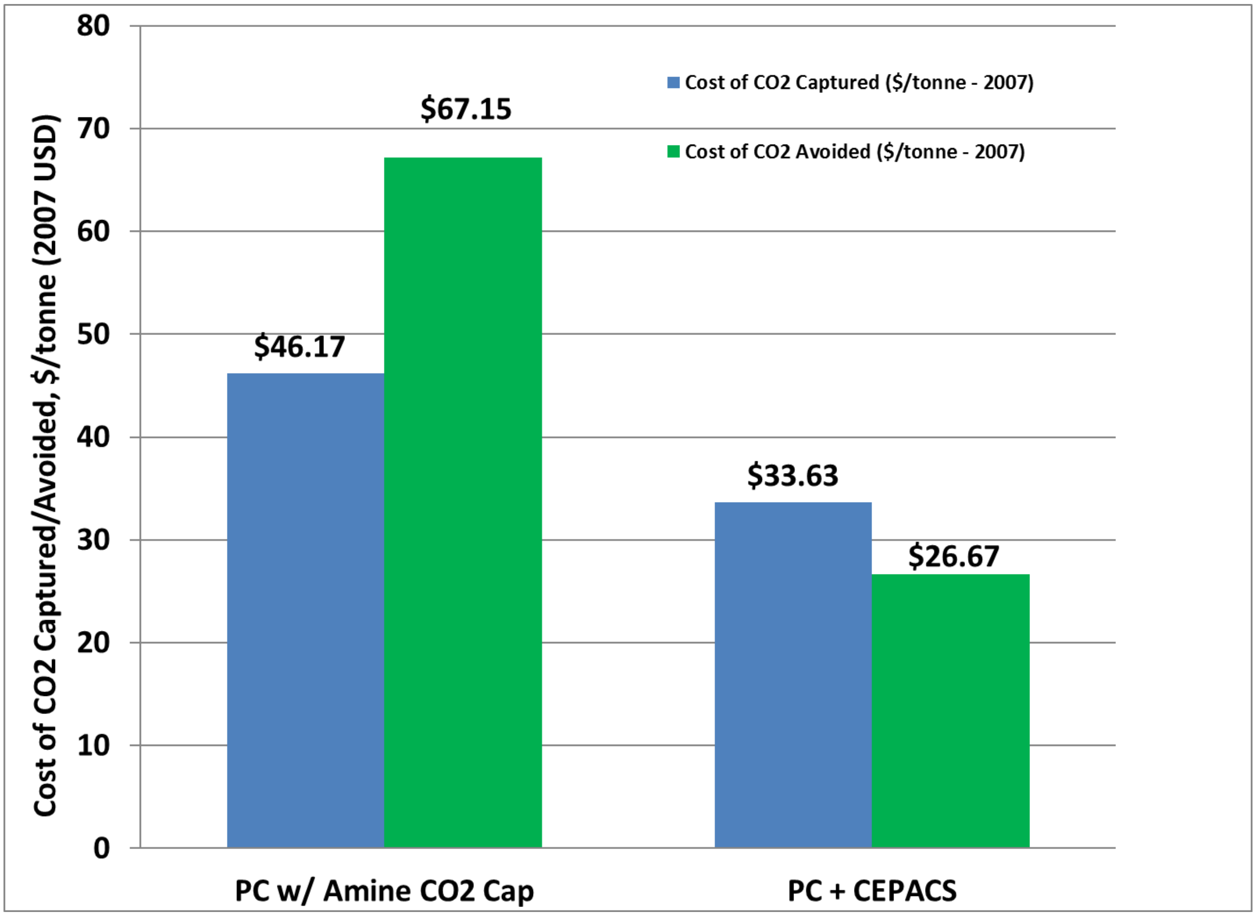


Figure 8-9. Cost of CO_2 Captured

Environmental Performance

Figure 8-10 shows the normalized emission rates of NOx, SO_2, and particulate matter (PM) calculated for each case. Emission limits from the US EPA new source performance standards (NSPS) Mercury and Air Toxics Standards (MATS) are also shown in the Figure. The following key conclusions can be drawn:

- Cases with carbon capture (Cases 2 and 3) require SO_2 polishing equipment added to the baseline PC plant and, as such, they have very low SO_2 emissions.
- Case 3 has the lowest emission rates of SO_2, NOx, and PM of all the cases studied.
- Case 3 has significantly lower emissions of SO_2, NOx, and PM than required by the NSPS MATS rules. Emissions of NOx and PM for Case 2 exceed the MATS requirements.

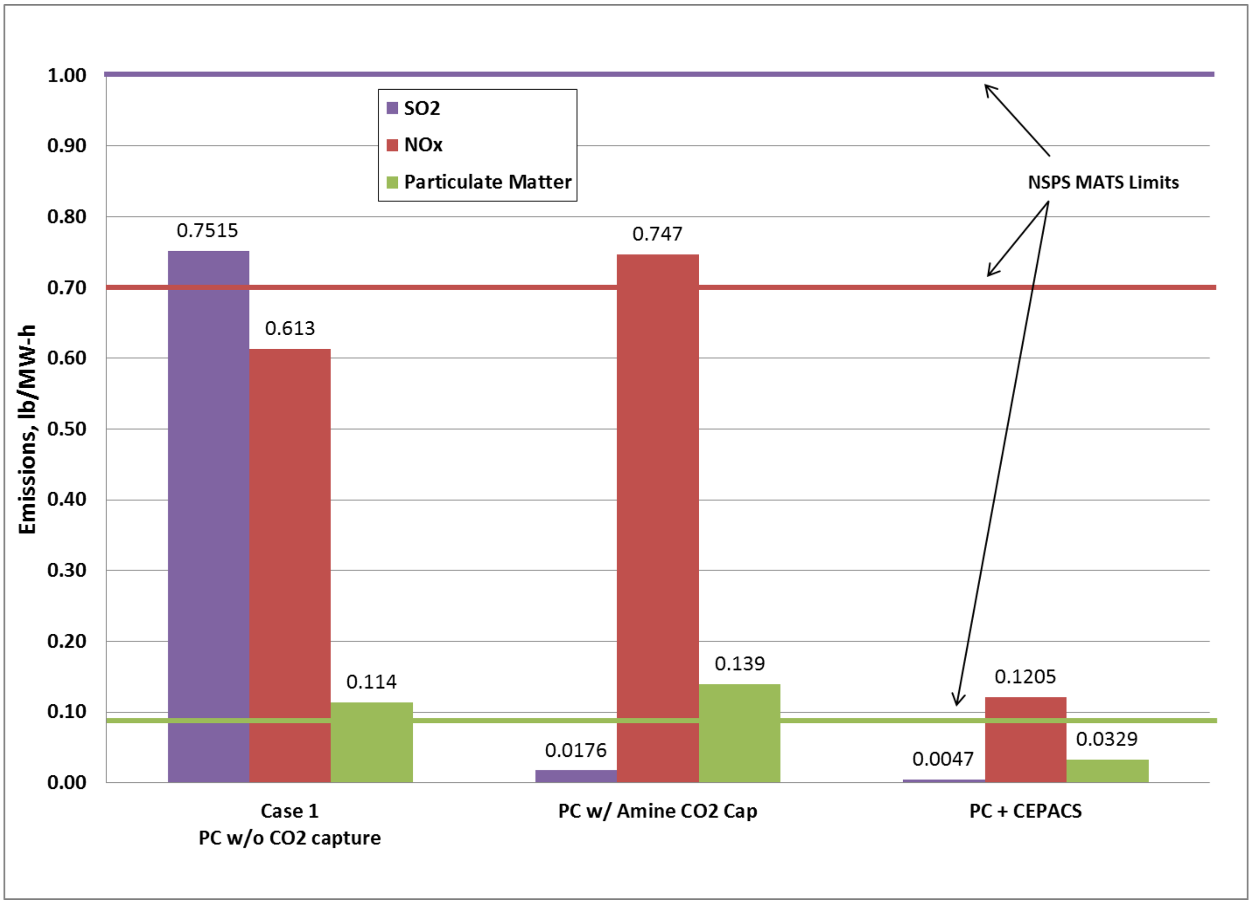


Figure 8-10. SO_2, NOx, and Particulate Emission Rates

Figure 8-11 shows the normalized emission rates of mercury calculated for each case. The following key conclusion can be drawn:

- Case 3 has lower Hg emissions than Cases 1 and 2 because a portion of the net power generation is fueled by natural gas, which contains negligible amounts of Hg. No additional Hg removal from the flue gas (due to the CEPACS system) was assumed for Case 3.

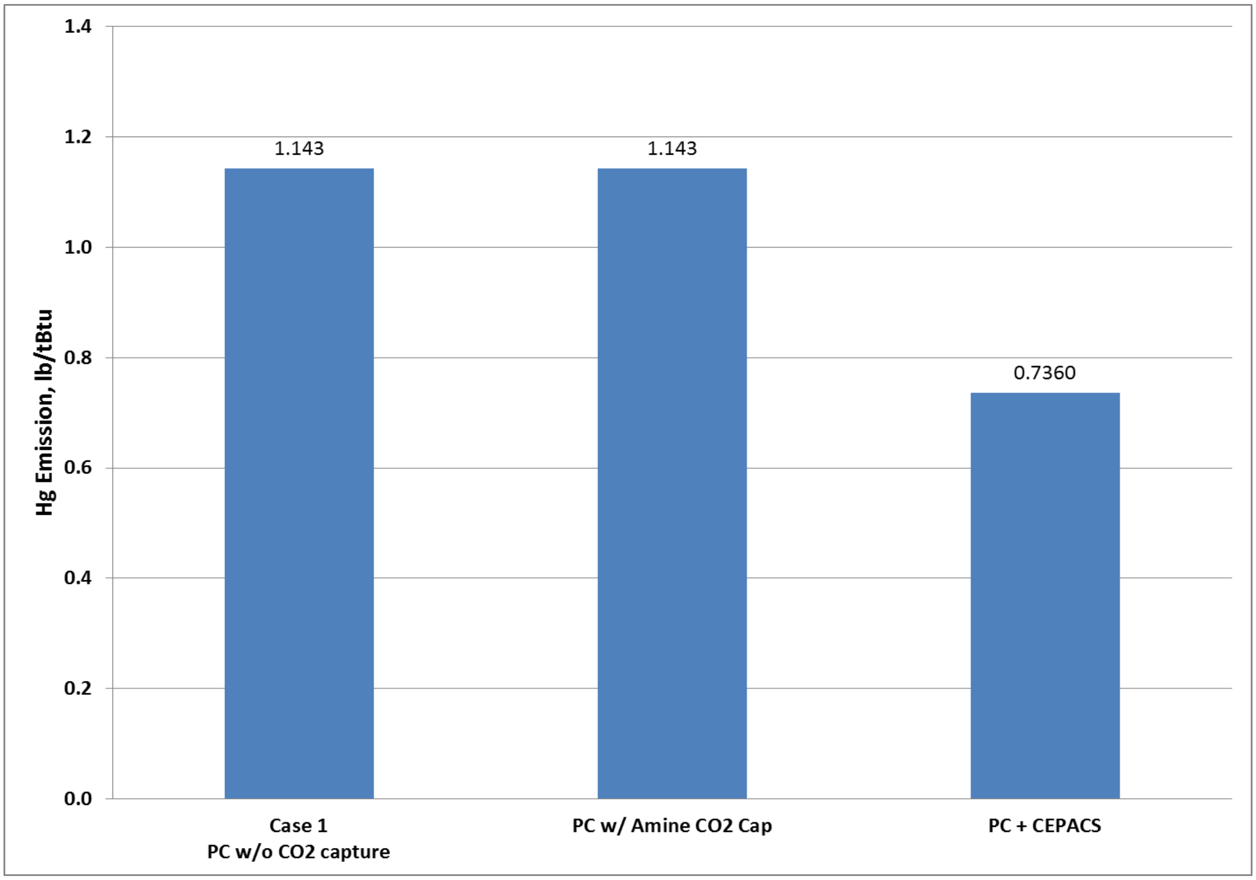


Figure 8-11. Mercury Emission Rates

Figure 8-12 shows the normalized emission rates of CO_2 calculated for each case. The following key conclusion can be drawn:

- Case 3 (PC + CEPACS) has the lowest CO_2 emissions of all the cases. This is due to the design criteria to capture 90% of the carbon which enters the CEPACS system. Since the CEPACS system generates additional power, the CO_2 emissions (normalized by total plant net output) are 58.3% lower than for Case 2.

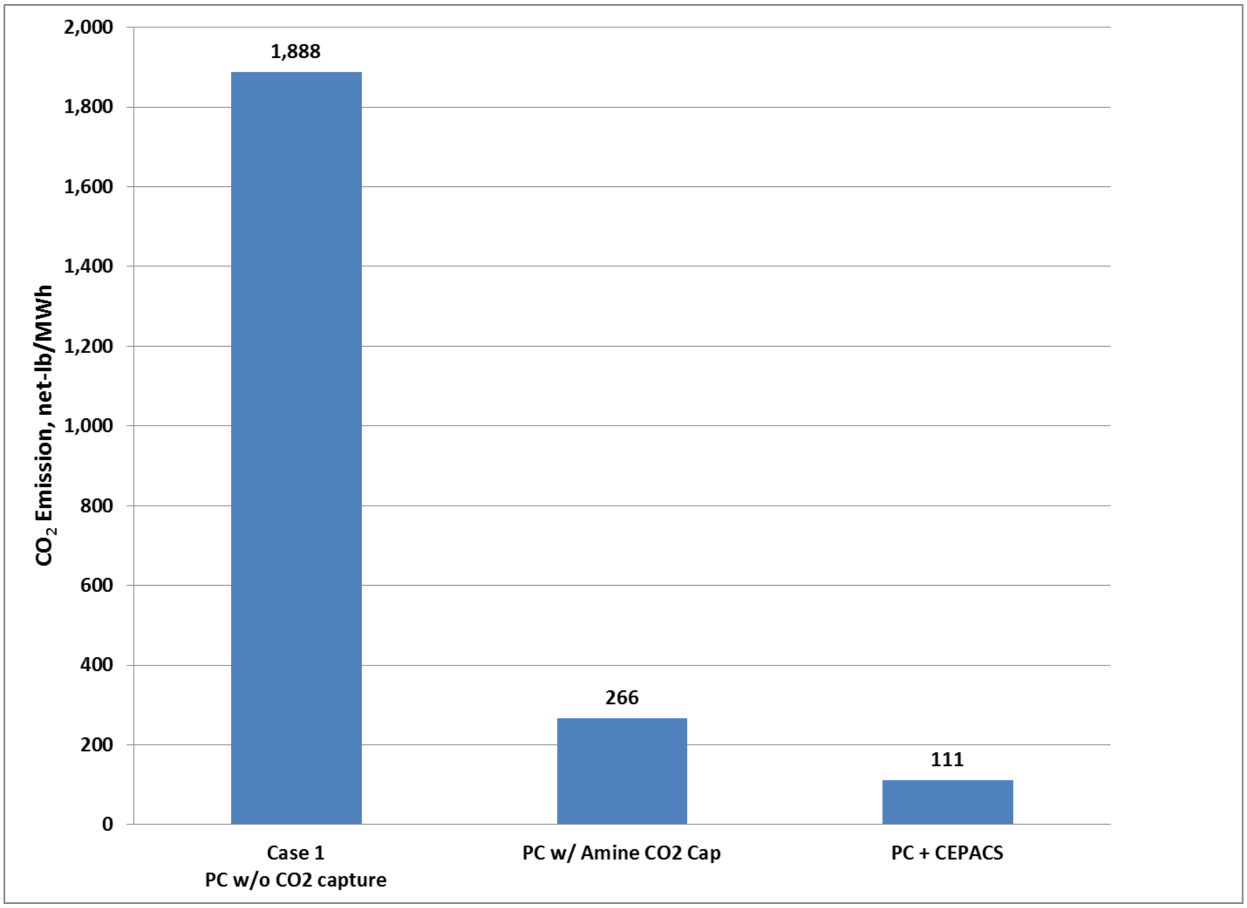


Figure 8-12. CO_2 Emission Rates

Overall, the CEPACS-based CO_2 capture process is more attractive technically and economically than the amine-based process.

CONCLUSION

Preliminary Technical and Economic Feasibility Study:
A CEPACS plant utilizing ECM technology was designed for a reference 550 MW (net AC) PC Rankine Cycle Plant to capture and compress CO_2 from the PC plant flue gas. The PC plant design specified in "Case 9" of the U.S. Department of Energy - National Energy Technology Laboratory (DOE - NETL) report "*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2" was* used as a reference. The performance assessment included estimation of the parasitic power consumption for >90% CO_2 capture and compression, and the efficiency impact on the PC plant. An equipment list and ECM module layout were prepared to facilitate the economic analysis. The CEPACS plant contained a total of 500 ECM modules. The economic feasibility study included estimation of CEPACS plant capital cost and cost of electricity (COE) analyses. While the ECM-based CEPACS system for the 550 MW PC plant captures 90% of CO_2 from the flue gas, it generates additional (net AC) power after compensating for the auxiliary power requirements of CO_2 capture and compression. The net electrical efficiency of the PC plant equipped with a CEPACS system (for CO_2 capture) was estimated to be 39.8% (based on higher heating values of coal and natural gas fuels used by PC plant and CEPACS plant, respectively). Case 3 (CEPACS-equipped PC plant) has the potential to meet DOE's objective

related to the incremental cost of electricity (COE) for post-combustion CO_2 capture (no more than 35% increase in COE). Case 3 also offers ~3 percentage point increase in PC plant efficiency as compared to the competing technology (Case 2, amine-based CO_2 capture) reducing the PC plant efficiency by more than 10 percentage points.

Flue Gas Contaminant Effect Evaluation:

The contaminant evaluation addressed possible interactions of the impurities that may be present in flue gas with ECM cell, which could result in reduced performance or life. Four main flue gas impurities were considered – sulfur, chlorine, mercury and selenium. The testing of the ECM cells was conducted by PNNL using FCE's button cells. Effect of the contaminants on ECM cell performance and endurance was studied. Long-term flue gas contaminant exposure tests were performed at a constant current density (i.e. constant CO_2 flux) in the range of 50-160 mA/cm^2, while monitoring the cell voltage. Tests with 0.4 - 1 ppmv SO_2 (in simulated PC plant flue gas used as cathode feed) included a cell operated at a constant CO_2 flux of 176 scc/m^2/s. To simulate an upset in the Flue Gas Desulfurization polishing unit of a CEPACS system, higher concentration levels of up to 40 ppm SO_2 were evaluated. While performance losses were observed on introduction of SO_2 at 10 ppm and higher levels, the voltage loss was found to be nearly fully reversible on return to a lower concentration level (1 ppm). Overall, tests showed negligible losses at the system design conditions of ~0.4 ppm SO_2 (in the effluent of the flue gas polisher in a CEPACS system). Button cell tests also included tests with HCl (in ECM cathode feed gas). Testing at concentration levels typical of coal power plants was performed. With 0.2 ppm HCl in flue gas, no measurable cell degradation (at 80 and 160 mA/cm^2) was observed during a 915-hour experiment. No performance losses were observed in tests of up to 1100 hours (at 110 mA/cm^2 and 650°C) due to the presence of 250 ppb Hg(g) in flue gas and up to 750 hours with 250 ppb Hg(g)+0.2 ppm HCl. No performance degradation was observed over an 600-hour test (at 160 mA/cm^2) with 10 ppb SeO_2 in flue gas.

Testing was performed at FCE using a larger 250 cm^2 ECM cell to characterize the ECM's ability to reduce NO_x emissions from the PC plant flue gas. NO_x (as NO) concentration in cathode feed stream (flue gas) was increased from 50 to 216 ppm. At least 70% of the NO_x was removed at 110 mA/cm^2 for the concentration levels tested. Testing also included analysis of the anode exit stream, which confirmed that NO_x was not transferred to the anode side. It was further observed that NO_x had no significant effect on the cell performance at the concentrations tested.

Based on the experimental results, contaminant tolerance levels for the ECM were identified. The contaminant levels expected from the flue gas clean-up (polishing) subsystem were estimated and compared with the ECM tolerance levels. The contaminant evaluation and comparison with CEPACS plant flue gas polishing system output showed that the ECM tolerance levels are well above the contaminant levels expected in the ECM cathode feed gas (treated flue gas).

Membrane Performance Characterization:

The membrane performance characterization tests were conducted on a 250 cm^2 ECM cell. The testing included operation at various current densities and reactant utilizations (gas feed flow rates). Cell performance (measured in volts), CO_2 flux (measured as current) and the % CO_2 transferred were characterized. The fuel utilization was varied from 30% to 68%, the CO_2 utilization (%CO_2 separated) was varied from 40% to 92%, and the current density was varied from 29.6 mA/cm^2 to 149.2 mA/cm^2. Cell temperature of 620°C and cathode inlet O_2 concentration of 8.3% (and CO_2 concentrations of 9.9%) were used as the baseline conditions for testing. From these test data, constant flow polarization and constant CO_2 (and fuel) utilization polarization curves were generated for the cell. After the characterization at baseline conditions, cell performance at various cathode inlet oxygen concentrations was examined. The cathode inlet oxygen concentration level was varied from 7% to 10% (representing the extent of air supplementation of the flue gas). From the parametric test data collected, the power production and CO_2 flux were calculated. As the current density (corresponds to CO_2 flux and hence % of cathode gas CO_2 separated for a constant cathode feed flow rate) increases, the power produced increases. The results showed that the cell performance increased as O_2 concentration increased up to 8.5% level and then began to decrease. This may be due to decreased CO_2 (the other oxidant needed for ECM operation) concentration resulting from increased air supplementation (CO_2 concentration decreases from 11.0 to 8.6% as O_2 concentration is increased from 7.0 to 10.0% in the test). At the baseline system operating conditions of 110 mA/cm^2, 68% fuel utilization, and 92% CO_2 utilization (percent CO_2 separated), a carbon dioxide flux rate of 128 scc/s/m^2 and a complete selectivity towards CO_2 transfer from the cathode to the anode were observed. All this testing was performed up to a fuel utilization of 68%. System simulation results showed improved overall system efficiency at a fuel utilization of 75%. Therefore, an evaluation of the effect of fuel utilization (at 110 mA/cm^2 and 93% CO_2 utilization) on cell performance was performed. The change from 68% to 75% fuel utilization resulted in a 2% decrease in cell voltage. The ECM cell performance was characterized at 75% fuel utilization (93% CO_2 utilization and 8.3% O_2) at a range of current densities. The constant utilization polarization characteristics were generated to guide selection of the ECM operating current density for the CEPACS system.

Balance of Plant Component Technology Gap Evaluation:
BOP equipment and related technology were evaluated to check for commercial availability or need for a custom design. Technical and cost quotations for key equipment were solicited. Technology development, if required, and the extent of development to fill the technology gap were identified as an outcome of the evaluation process. Major components for which the status of technology was examined included Flue Gas Clean-up subsystem, chillers, ECM stack enclosures and interconnections, and CO_2 compressors. Flue gas clean-up technologies are commercially available and require moderate development for this application. Traditional wet scrubber flue gas desulfurization (FGD) systems are available from many large globally-known OEMs (original equipment manufacturers). Key equipment in the CEPACS system for separating CO_2 from other gases present in the CO_2 – rich stream leaving the ECM anodes is the chillers. A survey of vendors indicated that chillers are available from a large number of suppliers for applications requiring chiller capacity up to about 5000 RT (refrigeration tons). Larger capacity chillers (up to 18,000 RT) are needed for the CEPACS system. A single train (e.g. centralized) chiller system was configured. The design resulted in two stages of chillers at two temperature levels. The first stage chiller design (moderate temperature) resulted in a single 4,400 RT chiller, well-aligned with the size range of the vendor's commercial products (up to 5,000 RT). For the second stage of chillers (low temperature), which required 18,000 RT (too large to supply as a single unit), the vendor quoted a modular design of five 3,600 RT units based on existing commercial offerings. The estimated chiller costs represented only ~3% of the Total Plant Cost, indicating economic viability for the process. Vendor contacts were made for compressors (CO_2-rich gas). FCE received multiple proposals from large, globally-known OEMs. The compressors in these proposals comprised state-of-the-art machinery, with a broad range of applications. The efficiencies ranged from 78.5 to 89.5%. The compressors in all of the proposals are suitable for operation at the CEPACS system process conditions. Overall, CO_2 compressors are commercially available and require no development for this application. As a part of the technology gap analysis, the CEPACS plant piping and ducting including the ECM module interconnections were also evaluated. In the plant layout developed, high (operating) temperature BOP equipment was located proximate to the ECM stack enclosures. Low temperature BOP equipment was centralized in a single area. This approach greatly reduced

the lengths of high temperature piping and ducting. The low temperature piping and ducting (including the headers) was designed so that only standard, commercially-off-the-shelf products were used. Vendors familiar with fabricating high temperature ducting for the Power Generation and Chemical Process Industries were identified.

Bench-scale ECM Stack and CEPACS System Test Demonstration:

A bench-scale system was designed and built to conduct demonstration testing to prove the feasibility of using ECM technology for CO_2 capture. The system utilized an ECM stack containing cells with a total electrochemical membrane area of 11.7 m^2. Construction of the CEPACS skid was completed. The test stack included 14 full-area cells. The cell assemblies were obtained from FCE's Torrington, CT commercial fuel cell manufacturing plant. Bench-scale testing of ECM-based CO_2 capture system was completed. The ECM stack was operated at CEPACS system operating conditions using simulated PC plant flue gas. The total test period of >15,700 h included steady state testing as well as the parametric and optimization tests. The stack completed the planned nine months of steady state testing, meeting one of the project milestones. The CO_2 flux of 116 cc/s/m^2 was maintained constant for over 6,500 hours, transferring more than 90% of the CO_2 in the cathode stream to the anode stream. The project technical milestone of verifying CO_2 flux at 100 cc/s/m^2 was achieved, as the CO_2 flux observed in the bench-scale test exceeded the required flux. The ECM stack had a constant gross DC power output of nearly 8 kW (>10 kW peak power), and experienced a power degradation rate of 0.05%/1000 h. The stack transferred more than 120 tonnes (metric tons) of CO_2 from the simulated flue gas to the anode exhaust stream over the total test period. Parametric testing, including operation at various CO_2 capture percentages and current densities (CO_2 flux), was conducted. The testing showed that the ECM stack is capable of operating at 20% higher CO_2 flux (~140 cc/s/m^2), with a proportional increase in power output. The required three thermal cycles were completed. There was a minimal performance penalty, in terms of DC power produced and no performance penalty in terms of CO_2 flux or current during the thermal cycling evaluation.

Testing of the CO_2 compression/liquefaction skid was carried out in parallel (with the bench-scale ECM stack CO_2 capture tests). The skid was evaluated under simulated ECM stack exhaust conditions, demonstrating the effective liquefaction and separation of high-purity CO_2. Additional testing included optimizing the operation of key system components. Several major advancements occurred in both the physical system hardware and system testing. System reliability upgrades were completed and any system deficiencies limiting extended duration testing were addressed. Following the system upgrades, a variety of operating conditions were applied to the CO_2 purification skid to characterize the system response, product characteristics and validate the operation. The results in general indicated the ability to produce food grade liquid CO_2 (>99.9% CO_2 and no measureable CO) across all tested operable points. Testing was transitioned from system development stage to system prototype maturation, operator training and system optimization for long term data collection processes. A key outcome was the training and validation of several system operators who were not involved with the development of the system hardware. This transition demonstrated the maturity of the CO_2 capture and purification system and its ability to be used as a parametric analysis tool rather than a development prototype.

Updated (Final) Technical and Economic Feasibility Study:

The final updated Technical & Economic Feasibility Study was conducted in BP3 of the project to assess the performance and cost of FCE's ECM-based CO_2 capture system. The CEPACS plant utilizing ECM technology was designed to capture and compress >90% of CO_2 from the flue gas of the reference 550 MW (net AC) Pulverized Coal (PC) Subcritical Steam Cycle Plant. Process simulations were updated based on ECM performance realized in bench-scale testing.

The updated CEPACS plant contained a total of 1,792 ECM stacks. The economic feasibility study included estimation of CEPACS plant capital cost, cost of electricity (COE) analyses and, estimation of cost per ton of CO_2 captured and cost per ton of CO_2 avoided. The CEPACS system employs 1,792 ECM stacks divided into 8 sections. For large-scale applications of the CEPACS system, grouping the stacks into larger enclosures reduces the overall capital costs by eliminating smaller module enclosures and their associated piping, facilitates the replacement of individual stacks (compared to replacing complete modules) and provides economies of scale that are not possible with 448 separate 1 MW (nominal), 4-stack modules (BP1 stack packaging concept). In each section, 224 ECM stacks are located in an enclosure. The layout of the CEPACS system was designed to minimize capital costs. The modular design minimized the lengths of “hot” piping and the quantities of fittings, significantly reducing capital costs while simplifying the sparing of parts and potentially increasing the capacity factor.

In the process of capturing ≥90% of the CO_2 from the PC plant flue gas, the ECM-based CEPACS system for the 550 MW (net AC) PC plant simultaneously generates 351 MW additional (net AC) power after compensating for the auxiliary power requirements of CO_2 capture and compression. The net electrical efficiency of the CEPACS-equipped PC plant (with CO_2 capture) was estimated to be 38.8% (based on higher heating values of coal and natural gas fuels used by PC plant and CEPACS system, respectively). The TOC of the subcritical PC power plant with the CEPACS system for CO_2 capture was estimated to be \$2,297/kW. The estimated COE for Case 3 is 80.9 mills/kWh. The incremental COE for Case 3 is 36.2%. The cost of CO_2 captured was calculated for the two CO_2 capture cases. The costs are shown in year 2007 US dollars and per metric ton of CO_2, for consistency with the T&EFS basis. The CEPACS system cost of CO_2 captured is \$33.63/tonne (2007 USD), which compares favorably with the DOE target of less than \$40/tonne (2011 USD). The cost of CO_2 captured for the CEPACS-based process (Case 3) is 27.2% lower than that for the Econamine-based process (Case 2). The cost of CO_2 captured for the amine-based process does not meet the DOE target. Overall, the CEPACS-based CO_2 capture process is more attractive technically and economically than the amine-based process.

REFERENCES

[1] Farooque, M., “Integrated high efficiency fossil fuel power plant/fuel cell system with CO2 emissions abatement”, United States Patent 7,396,603, issued July 08, 2008.

[2] Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2, DOE/NETL-2010/1397, November 2010.

[3] Quality Guidelines for Energy System Studies: Cost Estimation Methodology for NETL Assessment of Power Plant Performance, April 2011, DOE/NETL-2011/1455

[4] Scope and Reporting Requirements for NETL System Studies, May 2010, DOE/NETL

[5] R. Ciccoli, V. Cigolotti, R. Lo Presti, E. Massi, S.J. McPhail, G. Monteleone, A. Moreno, V. Naticchioni, C. Paoletti, E. Simonetti, F. Zaza, Molten carbonate fuel cells fed with biogas: Combating H2S, Waste Management, 30 (2010) 1018-1024.

[6] T.D. Claar, E.H. Camara, R.A. Donado, L.G. Marianowski, MOLTEN CARBONATE FUEL-CELL BEHAVIOR WITH SULFUR-CONTAINING GASES, Journal of the Electrochemical Society, 125 (1978) C471-C471.

[7] H. Devianto, E. Simonetti, S.J. McPhail, F. Zaza, V. Cigolotti, C. Paoletti, A. Moreno, A. La Barbera, I. Luisetto, Electrochemical impedance study of the poisoning behavior of Ni-based anodes at low concentrations of H2S in an MCFC, International Journal of Hydrogen Energy, 37 (2012) 19312-19318.

[8] H. Devianto, S.P. Yoon, S.W. Nam, J. Han, T.H. Lim, The effect of a ceria coating on the H2S tolerance of a molten carbonate fuel cell, Journal of Power Sources, 159 (2006) 1147-1152.

[9] M. Kawase, Y. Mugikura, Y. Izaki, T. Watanabe, Effects of H2S on the performance of MCFC II. Behavior of sulfur in the cell, Electrochemistry, 67 (1999) 364-371.

[10] M. Kawase, Y. Mugikura, Y. Izaki, T. Watanabe, J. Fukuda, Effects of H2S on the performance of MCFC - I. Change of cell voltage and gas composition, Denki Kagaku, 66 (1998) 449-455.

[11] G. Monteleone, M. De Francesco, S. Galli, M. Marchetti, V. Naticchioni, Deep H2S removal from biogas for molten carbonate fuel cell (MCFC) systems, Chemical Engineering Journal, 173 (2011) 407-414.

[12] A.F. Sammells, S.B. Nicholson, P. Ang, DEVELOPMENT OF SULFUR-TOLERANT COMPONENTS FOR MOLTEN CARBONATE FUEL-CELL, Journal of the Electrochemical Society, 125 (1978) C373-C373.

[13] A.F. Sammells, S.B. Nicholson, P.G.P. Ang, DEVELOPMENT OF SULFUR-TOLERANT COMPONENTS FOR THE MOLTEN CARBONATE FUEL-CELL, Journal of the Electrochemical Society, 127 (1980) 350-357.

[14] S.W. Smith, H.R. Kunz, W.M. Vogel, S.J. Szymanski, EFFECTS OF SULFUR ON MOLTEN CARBONATE FUEL-CELLS, Journal of the Electrochemical Society, 129 (1982) C118-C118.

[15] F. Zaza, C. Paoletti, R. LoPresti, E. Simonetti, M. Pasquali, Studies on sulfur poisoning and development of advanced anodic materials for waste-to-energy fuel cells applications, Journal of Power Sources, 195 (2010) 4043-4050.

[16] S. Chamberlain, T. Reeder, C.K. Stimpson, D.R. Tree, A comparison of sulfur and chlorine gas species in pulverized-coal, air- and oxy-combustion, Combustion and Flame, 160 (2013) 2529-2539.

[17] N. Tsubouchi, T. Saito, N. Ohtaka, Y. Ohtsuka, Evolution of Hydrogen Chloride and Change in the Chlorine Functionality during Pyrolysis of Argonne Premium Coal Samples, Energy & Fuels, 27 (2013) 87-96.

[18] R. Meij, B.H.T. Winkel, Trace elements in world steam coal and their behavior in Dutch coal-fired power stations: A review, International Journal of Coal Geology, 77 (2009) 289-293.

[19] R. Zevenhoven, P. Kilpinen, Chapter 7. Halogens, dioxins/furans, in: Control of Pollutants in Flue Gases and Fuel Gases, Helsinki University of Technology, Espoo/Turku, Finland, 2002.

[20] R. Yan, D. Gauthier, G. Flamant, Volatility and chemistry of trace elements in a coal combustor, Fuel, 80 (2001) 2217-2226.

[21] R. Yan, D. Gauthier, G. Flamant, Possible interactions between As, Se, and Hg during coal combustion, Combustion and Flame, 120 (2000) 49-60.

[22] R. Yan, D. Gauthier, G. Flamant, G. Peraudeau, Fate of selenium in coal combustion: Volatilization and speciation in the flue gas, Environmental Science & Technology, 35 (2001) 1406-1410.

[23] R. Yan, D. Gauthier, G. Flamant, Y. Wang, Behavior of selenium in the combustion of coal or coke spiked with Se, Combustion and Flame, 138 (2004) 20-29.

[24] O.A. Marina, L.R. Pederson, C.A. Coyle, E.C. Thomsen, D.J. Edwards, Polarization-Induced Interfacial Reactions Between Nickel and Selenium in Ni/Zirconia SOFC Anodes and Comparison with Sulfur Poisoning, Journal of the Electrochemical Society, 158 (2011) B36-B43.

[25] O.A. Marina, L.R. Pederson, R. Gemmen, K. Gerdes, H. Finklea, I.B. Celik, Overview of SOFC Anode Interactions with Coal Gas Impurities, in: M.C. Williams, K. Krist, N. Garland (Eds.) Fuel Cell Seminar 2009, 2010, pp. 363-370.

[26] Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity', Revision 2a, DOE/NETL-2010/1397, September 2013

LIST OF ACRONYMS

AC	Alternating Current
ACFM	Actual Cubic Feet per Minute
AECOM	AECOM Corporation (Formerly URS Corporation)
AES	Auger Electron Spectroscopy
BEC	Bare Erected Cost
BFD	Block Flow Diagram
BFW	Boiler Feed Water
BOP	Balance of Plant
BP	Budget Period
CCF	Capital Charge Factor
CCUS	Carbon Capture, Use, and Sequestration
CRADA	Cooperative Research and Development Agreement
CEPACS	Combined Electric Power And Carbon-dioxide Separation
CF	Capacity Factor
COE	Cost of Electricity
CP	Coefficient of Performance
DC	Direct Current
DCS	Distributed Control System
DFC	Direct Fuel Cell (FCE Molten Carbonate Fuel Cell)
DOE	United States Department of Energy
EBSD	Electron Back-Scatter Diffraction
ECM	Electro-Chemical Membrane
EH&S	Environmental Health and Safety
EPC	Engineering, Procurement and Construction
EPCC	Engineering, Procurement and Construction Cost
FCE	FuelCell Energy, Inc.
FD	Forced Draft
FGD	Flue Gas Desulfurization
GHG	Greenhouse Gases
HHV	Higher Heating Value
H&MB	Heat and Mass Balance
HRSG	Heat Recovery Steam Generator
HSS	Heat Stable Salts

HVAC	Heating Ventilation Air Conditioning
IOU	Investor Owned Utility
IRROE	Internal Rate of Return On Equity
kV	kilo-Volt
kW, kWe, kWt	Kilo-Watt, Kilo-Watt electricity, Kilo-Watt thermal equivalent
LCOE	Levelized Cost of Electricity
LHV	Lower Heating Value
LTSC	Low Temperature Shift Converter
LTSR	Low Temperature Shift Reactor
MEA	Monoethanolamine
MW	Mega-Watt
NETL	National Energy Technology Laboratory
NG	Natural Gas
NGCC	Natural Gas Combined Cycle
OC_{FIX}	Fixed Operating Cost
OC_{VAR}	Variable Operating Cost
OEM	Original Equipment Manufacturer
OFA	Overfire Air
O&M	Operating and Maintenance
PC	Pulverized Coal
PFD	Process Flow Diagram
PM	Project Manager
PMP	Project Management Plan
PNNL	Pacific Northwest National Laboratory
PT&EFS	Preliminary Technical & Economic Feasibility Study
ppm	parts per million
psia	Pound per square inch absolute pressure
RMM	Risk Management Methodology
RU	Reforming Unit
SCFM	Standard Cubic Feet per Minute
SCR	Selective Catalytic Reduction
SEM	Scanning Electron microscopy
SOPO	Statement of Project Objectives
TASC	Total As-spent Cost
TEM	Transmission Electron microscopy

TOC	Total Overnight Cost
TS&M	Transport, Storage & Monitoring
ToF-SIMS	Time-of-Flight Secondary Ion Mass Spectroscopy
TPC	Total Plant Cost
US	United States of America
USD	US Dollars
XPS	X-ray Photoelectron Spectroscopy
XRD	X-Ray Diffraction