Evaluation of Solid Sorbents as a Retrofit Technology for CO_2 Capture

Final Technical Report

DOE Award DE-FE004343

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Abstract

ADA completed a DOE-sponsored program titled Evaluation of Solid Sorbents as a Retrofit Technology for CO_2 Capture under program DE-FE0004343. During this program, sorbents were analyzed for use in a post-combustion CO_2 capture process.

A supported amine sorbent was selected based upon superior performance to adsorb a greater amount of CO_2 than the activated carbon sorbents tested. When the most ideal sorbent at the time was selected, it was characterized and used to create a preliminary techno-economic analysis (TEA). A preliminary 550 MW coal-fired power plant using Illinois #6 bituminous coal was designed with a solid sorbent CO_2 capture system using the selected supported amine sorbent to both facilitate the TEA and to create the necessary framework to scale down the design to a 1 MWe equivalent slipstream pilot facility.

The preliminary techno-economic analysis showed promising results and potential for improved performance for CO_2 capture compared to conventional MEA systems. As a result, a 1 MWe equivalent solid sorbent system was designed, constructed, and then installed at a coal-fired power plant in Alabama.

The pilot was designed to capture 90% of the CO_2 from the incoming flue gas at 1 MWe net electrical generating equivalent. Testing was not possible at the design conditions due to changes in sorbent handling characteristics at post-regenerator temperatures that were not properly incorporated into the pilot design. Thus, severe pluggage occurred at nominally 60% of the design sorbent circulation rate with heated sorbent, although no handling issues were noted when the system was operated prior to bringing the regenerator to operating temperature. Testing within the constraints of the pilot plant resulted in 90% capture of the incoming CO_2 at a flow rate equivalent of 0.2 to 0.25 MWe net electrical generating equivalent. The reduction in equivalent flow rate at 90% capture was primarily the result of sorbent circulation limitations at operating temperatures combined with pre-loading of the sorbent with CO_2 prior to entering the adsorber. Specifically, CO_2-rich gas was utilized to convey sorbent from the regenerator to the adsorber. This gas was nominally 45°C below the regenerator temperature during testing.

ADA's post-combustion capture system with modifications to overcome pilot constraints, in conjunction with incorporating a sorbent with CO_2 working capacity of 15 g CO_2/100 g sorbent and a contact time of 10 to 15 minutes or less with flue gas could provide significant cost and performance benefits when compared to an MEA system.

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List of Acronyms and Abbreviations

APTES	3-(aminopropyl)triethoxysilane
°C	degree Celsius
cm	centimeter
CO	Colorado
CO_2	carbon dioxide
DOE	United States Department of Energy
DSC	differential scanning calorimeter
EOR	enhanced oil recovery
ESP	electrostatic precipitator
°F	degree Fahrenheit
FAI	FAI Materials Testing Laboratory
FCC	fluidized catalytic cracking
FF	fabric filter
FGD	flue-gas desulfurization
ft	feet
g	gram
H_2O	water
HHV	higher heating value
hr	hour
ID	inner diameter
KJ	Kilojoule
KW	kilowatt
L	liter
lb	pound
m	meter
MEA	monoethanolamine
mg	milligram
min	minute
mL	milliliter
mm	millimeter
MOF	metal organic frameworks
mol	mole
MS	mass spectrometer
MW	megawatt
NETL	National Energy Technology Laboratory
N_2	nitrogen
PC	post-combustion
PCO_2	partial pressure of carbon dioxide
PEI	polyethlyeneimine
sec	second
SO_2	sulfur dioxide
TGA	thermogravimetric analysis
TSA	temperature-swing adsorption

1. Executive Summary

Carbon dioxide capture and storage (CCS) is anticipated to be an important component of reducing worldwide CO_2 emissions from stationary point sources, such as coal-fired power plants. One of the most important challenges to the widespread implementation of CCS is the cost and energy associated with the separation of CO_2 from flue gas. Aqueous amines and ammonia are being demonstrated by several groups for CO_2 capture in a temperature-swing cyclic process. Solid sorbents can also be used in a similar temperature-swing process, but have the potential to reduce the overall costs related to CO_2 capture by reducing the energy required to release the CO_2 during material regeneration due to less evaporation of water and lower specific heat capacity. From 4th quarter 2010 through 3rd quarter 2015, ADA completed a DOE-sponsored project titled Evaluation of Solid Sorbents as a Retrofit Technology for CO_2 Capture under program DE-FE0004343. The objective of the project was to validate solid sorbent-based post-combustion CO_2 capture through slipstream pilot testing. During the project, sorbents were analyzed to select a sorbent for testing, a process design was finalized, and a pilot was designed, fabricated, and used for testing.

Both supported amines and activated carbon sorbents were considered and evaluated during this project. A supported amine sorbent was selected based upon superior performance to adsorb a greater amount of CO_2 than the activated carbon sorbents tested. When the most ideal sorbent at the time was selected, it was characterized and used to create a preliminary techno-economic analysis (TEA). A preliminary 550 MW coal-fired power plant using Illinois #6 bituminous coal was designed with a solid sorbent CO_2 capture system using the selected supported amine sorbent to both facilitate the TEA and to create the necessary framework to scale down the design to a 1 MWe equivalent slipstream pilot facility.

The preliminary techno-economic analysis showed promising results and potential for improved performance for CO_2 capture compared to conventional MEA systems. As a result, a 1 MWe equivalent solid sorbent system was designed, constructed, and then installed at a coal-fired power plant in Alabama.

The pilot was designed to capture 90% of the CO_2 from the incoming flue gas at 1 MWe net electrical generating equivalent. Testing was not possible at the design conditions due to changes in sorbent handling characteristics at post-regenerator temperatures that were not properly incorporated into the pilot design. Thus, severe pluggage occurred at nominally 60% of the design sorbent circulation rate with heated sorbent, although no handling issues were noted when the system was operated prior to bringing the regenerator to operating temperature. Testing within the constraints of the pilot plant resulted in 90% capture of the incoming CO_2 at a flow rate equivalent of 0.2 to 0.25 MWe net electrical generating equivalent. The reduction in equivalent flow rate at 90% capture was primarily the result of sorbent circulation limitations at operating temperatures combined with pre-loading of the sorbent with CO_2 prior to entering the adsorber. Specifically, CO_2-rich gas was utilized to convey sorbent from the regenerator to the adsorber. This gas was nominally 45°C below the regenerator temperature during testing.

A second TEA was performed to account for current limitations of the pilot. As expected, unless modifications are made to accommodate changes in sorbent handling characteristics at regeneration temperature and to limit pre-loading of the sorbent with CO_2 prior to the adsorber, the levelized cost of electricity (LCOE) will be significantly impacted. Furthermore, utilizing the system with modifications to overcome pilot constraints, in conjunction with incorporating a sorbent with CO_2 working capacity of 15 g CO_2/100 g sorbent and a contact time of 10 to 15 minutes or less with flue gas could provide significant cost and performance benefits when compared to an MEA system.

1. Introduction

Capturing and geologically storing the carbon dioxide (CO_2) from point source emitters may be one of the best options for controlling anthropogenic CO_2 emissions. More specifically, post-combustion CO_2 capture and sequestration is one of the only feasible means to significantly reduce CO_2 emissions from existing fossil fuel power plants. Although no post-combustion capture technology had been demonstrated at full-scale applications (greater than 25 MW) for coal-fired utilities at the time of project initiation, the most developed process utilized an aqueous amine solvent, often monoethanolamine (MEA), to react with the CO_2 at low temperature and then release it in a purified form at a higher temperature. The energy penalty associated with solvent-based temperature-swing processes is relatively high. For example, studies have shown that aqueous MEA for 90% CO_2 capture from a retrofit coal-fired power plant can reduce the thermal efficiency from approximately 35% (HHV basis) to 24.4% and cost \$80 per ton CO_2 removed.[1] Much of this cost is associated with the energy penalty, primarily heating and cooling the solvent, that is incurred when releasing the purified CO_2 in the heating/regeneration step.

The US Department of Energy National Energy Technology Laboratory has supported the development of alternative CO_2 capture technologies to reduce the overall costs and energy penalty associated with CO_2 mitigation from power generated using coal as the fuel source. With cost estimates of CO_2 capture technology as high as \$80 per ton CO_2 removed for first and second-generation systems, widespread post-combustion CO_2 capture would be extremely unlikely without advances to reduce costs.

Solid sorbents can also be used in a process that utilizes a temperature swing, referred to as temperature-swing adsorption (TSA). A temperature-swing process that utilizes solids has the potential to reduce the energy penalty associated with post-combustion CO_2 capture process. A simplified schematic of a TSA process is provided in Figure 1. The project reported on herein was designed to investigate whether solid sorbents used in a temperature-swing process for CO_2 capture could significantly reduce costs associated with post-combustion CO_2 capture. Under cooperative agreement with the Department of Energy (DE-FE0004343), ADA-ES, Inc. (ADA) led the development and testing of a sorbent-based CO_2 capture technology through process evaluation, 1 MWe pilot testing, and a techno-economic assessment during the project period of October 2010 through September 2015.

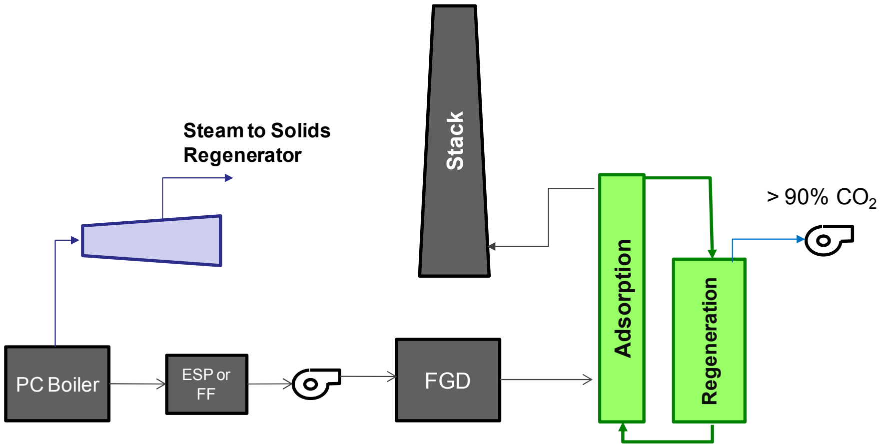

Figure 1: Generic Post-Combustion Temperature-swing Adsorption Process.

1.1 Overview of the Project

The overall objective of this project was to validate solid sorbent-based post-combustion CO_2 capture through slipstream pilot testing. Results from pilot testing were used to develop a preliminary full-scale commercial design. The DOE Program Goals, and the project goals at the outset of the program, were to achieve 90 percent CO_2 capture while limiting the increase in LCOE to 35 percent for an nth-of-a-kind full scale commercial system, with initial commercial deployment beginning in 2020.

1.1.1 Background and History of the Technology

While most solid sorbents offer the benefit of a lower specific heat compared to aqueous solutions, the mechanism in which such materials remove CO_2 from flue gas can vary significantly. Sorbents can be classified into two general families: those that chemically react with the CO_2, called supported reactants, and those that adsorb or use their molecular structure or Van der Waals forces to screen CO_2 from other gases, called non-reacting adsorbents. Chemically reacting sorbents usually include an inert, high surface area support, with an immobilized amine or other reactant on the surface. The surface area provides numerous sites for the desired reaction to occur. Many different types of solid materials for CO_2 capture have been or are currently being investigated including: supported amines[2-8], carbon-based sorbents[9-12], supported carbonates[13,14], zeolites[15], metal organic frameworks (MOFs)[16-20], etc. These materials are being developed and tested at universities, government laboratories, and private institutions worldwide. A great deal of research related to sorbent development and evaluation has occurred to date, but for the most part these promising materials had yet to be paired with a feasible process and demonstrated at the pilot scale.

Solid sorbent technology at the time of the award was in the early stages of development and required further research and demonstration before being considered a commercially viable option. One of the first steps in investigating CO_2 capture using solid sorbents was to determine which sorbent would be used for process design efforts. An extensive sorbent screening program was previously completed under cooperative agreement DE-NT0005649 as well as several other funding mechanisms including internal research. Based on the 140 sorbents evaluated during that program and more than 100 additional sorbents that were evaluated, it was determined that supported amine and carbon-based sorbents presented the best potential to significantly reduce the energy penalty associated with post-combustion CO_2 capture. However, such sorbents must be paired with a system/process that can take advantage of their beneficial properties.

Several process technology options were assessed during this program, including an entrained flow reactor, trickle-down reactor, moving-bed reactor, and fluidized bed reactor. In the absence of large scale (i.e. pilot-scale) testing, it was necessary to use laboratory results or make assumptions regarding the equipment and sorbent performance under the expected operating conditions. A fluidized-bed reactor was determined to be the most optimal for the sorbent chosen for the pilot program due to the optimal heat and mass transfer characteristics between gases and solids, and because very large fluidized beds were in use commercially in the petrochemical industry. A unique TSA process was developed during this project that incorporated a three-stage fluidized bed adsorber integrated with a single-stage fluidized bed regenerator.

Prior to this program, an initial techno-economic analysis was completed using a generic TSA process design to determine the potential of solid sorbents to reduce the costs associated with post-combustion CO_2 capture relative to a solvent-based process. The results of this preliminary assessment concluded:

- Projected capital costs, fuel costs and CO_2 transport, storage, and monitoring costs of the sorbent-based process were lower than those of the MEA case, respectively. Specifically, to generate 550 MWnet with 90% CO_2 capture, the gross power plant was smaller when sorbent-based CO_2 capture was used compared to MEA due to the lower thermal input required to operate the CO_2 capture facility.
- The projected operating and maintenance costs of the sorbent-based CO_2 capture were higher than those of the MEA process, mostly as a result of a high sorbent cost, the amount of sorbent required to initially fill the system, and sorbent replacement.
- The projected electric requirement for the compression of CO_2 was significant for both aqueous amine and sorbent-based CO_2 capture. However, because the gross power plant with sorbent-based capture was smaller, less CO_2 must be compressed and the related costs were accordingly less.
- The projected cost of electricity (COE) and levelized cost of electricity (LCOE) with sorbent-based CO_2 capture were calculated to be 113.3 and 143.6 mills/kWh, respectively. The projected COE and LCOE of the sorbent-based process were lower than those of the aqueous MEA system due to a lower capital costs, fuel costs, and CO_2 TS&M costs, offsetting higher O&M costs.

Although initial estimates of the costs of the sorbent-based CO_2 capture process resulted in a lower projected energy penalty and lower projected overall increase in the COE compared to a solvent-based process, the improvements with respect to an aqueous MEA process were not enough to meet the DOE's cost of CO_2 captured and cost of electricity goals. Therefore, this project was designed to make progress towards the overall DOE goals. Further improvements outside the scope of this project could provide additional progress towards the DOE goals. Potential areas of further improvement include:

- Heat integration
    - Heat recovery between the rich and lean sorbent – this must be accomplished in such a way that the energy recovery is significant enough to offset the added capital costs for the heat exchangers
    - Integration within the power plant – using heat from the power plant or the CO_2 compression system to reduce the overall energy penalty
- Reduce capital costs
- Improve sorbent performance and characteristics
    - Working CO_2 capacity
    - Faster kinetics
    - Attrition resistance
    - Cost
    - Thermal conductivity
    - Moisture loading
    - Etc.
- Optimize steam condition at IP/LP crossover to eliminate a need for the BP turbine and to integrate into the steam cycle in a more efficient manner.

In addition to potential cost savings, interest in solid sorbents was enhanced by environmental concern for competing solvent technology. In particular, a significant concern is the volatile emissions of amines. At the expected regeneration temperature the vapor pressure of 3.5 M aqueous MEA is approximately 0.0725 psi; emissions of amines could be a significant environmental concern for aqueous CO_2 capture systems. The sorbent selected for pilot-scale evaluation under this project consists of an ion exchange resin that incorporates amines that are covalently bonded to the substrate. Although it is possible to degrade the selected sorbent using excessively high temperatures, no measureable volatile emissions were recorded using a mass spectrometer. It is possible that the amine on the sorbent remains covalently attached to the surface of the substrate even when being converted to urea, heat stable salts, or being otherwise degraded. This is a beneficial characteristic for a sorbent used in a full-scale system.

Another environmental consideration is related to potential spills. If the sorbent used in a solid-sorbent-based process were to spill either in transport or at the power plant, neither the sorbent nor the amine will leach into the soil and the sorbent can be readily be cleaned up without serious environmental consequences. Environmental impacts due to inadvertent leaks are a concern for systems using aqueous solutions. The sorbent identified for pilot testing under this project is not considered hazardous by the OSHA Hazard Communication Standard, thus waste disposal should be in accordance with the plants locale environmental controls.

Finally, concerns exist that recycled aqueous solutions may concentrate heavy metals such as selenium or mercury leading to a potential hazardous waste issue. By using a solid sorbent rather than water-based solvents, water soluble contamination concentration may potentially be avoided.

1.1.2 Project Organization

ADA-ES was the prime recipient of award DE-FE0004343 and coordinated the efforts of multiple subcontractors. Project participants were selected based upon their qualifications and expertise as it relates to the scope of work assigned to them. ADA was responsible for oversight of pilot engineering design and construction. ADA led efforts of pilot operation, sorbent selection and scale up management, and quality assurance oversight. The project utilized project management procedures which required each subcontractor to submit monthly update reports. This information was used to monitor various project areas for each subcontractor: e.g., scope of work progress, cost, and schedule.

The project participants included:

Stantec Consulting Ltd.

ADA, along with Stantec, combined empirical data collected from prior field testing and laboratory experiments and used an iterative modeling approach to enhance the conceptual design effort. Stantec used the results from the modeling effort to evaluate a 500 MWe conceptual design which was used as the basis for the 1 MWe pilot design.

At the conclusion of testing, Stantec incorporated key findings from the testing of the pilot facility and conducted a techno-economic assessment of the commercial carbon capture system. Project costs were broken into four categories, as suggested through the DOE EPEC program.

Technip Stone and Webster Process Technology

Technip (formerly Shaw Energy & Chemicals, Inc.) conducted a technical review of the conceptual designs presented by ADA. ADA shared this information with Technip who further developed conceptual and final engineering and design details. Upon completion of the conceptual design efforts, preliminary cost projections for the 500 MWe commercial system were prepared by Stantec Consulting and reviewed.

Technip provided the technical engineering and design services for preparing the initial 1 MW pilot plant design. The pilot plant design was based upon the full-scale conceptual design developed for the 500 MWe retrofit application.

At the beginning of Budget Period 2, procurement activities commenced and detailed information for select equipment was developed to finalize fabrication and construction work packages. Technip used information from various equipment suppliers to finalize various connection and sizing details contained in the work packages. Technip had also been contracted to provide a fluidized bed expert to support commissioning and startup activities.

Electric Power Research Institute (EPRI)

EPRI is an independent membership-based non-profit company comprised primarily of electric power owners and operators that performs research, development and design in the electricity sector.

Southern Company

Southern company provided a host site for the 1 MWe pilot plant at one of its coal fired boilers located in Alabama. In addition, Southern Company provided cost share assistance to the project.

Luminant

Luminant provided cost share support for technology development.

1.1.3 Project Description

The overall objective of this program was to validate solid sorbent-based post-combustion CO_2 capture through slipstream pilot testing. Results from pilot testing were used to develop a preliminary full-scale commercial design. The work was completed in nine tasks and three phases, as shown in Table 1.

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Table 1. Project Task List and Budget Period.

Task	Phase I: Design Budget Period 1	Phase II: Build Budget Period 2	Phase III: Test Budget Period 3
Task 1: Project Management and Planning	X	X	X
Task 2: Refine Full-Scale Conceptual Design and Sorbent Selection	X		
Task 3: Design Pilot Equipment	X		
Task 4: Procure & Manufacture Sorbents		X	
Task 5: Procure and Construct Pilot-Scale Equipment		X	
Task 6: Install/Start-Up Pilot-Scale Equipment		X	
Task 7: Pilot-Scale Operation and Evaluation			X
Task 8: Define and Collect Compression and Sequestration-Specific Information			X
Task 9: Prepare Commercial Conceptual Design and Economics			X

1.1.4 Project Location

The project took place in various locations throughout North America. ADA's primary laboratory, project management, and engineering work was performed at ADA's headquarters in Highlands Ranch, CO. The ADA-led field testing and pilot commissioning occurred just north of Birmingham, AL at a power plant site.

Engineering and commercial design work was performed in Boston, MA. Techno-economic evaluations and support work were conducted in Regina, Saskatchewan Canada. Fabrication and Installation was conducted in multiple locations in Alabama.

1.1.5 Project Schedule

The project schedule seen in Figure 2: Project Milestone Schedule, was developed and managed by ADA.

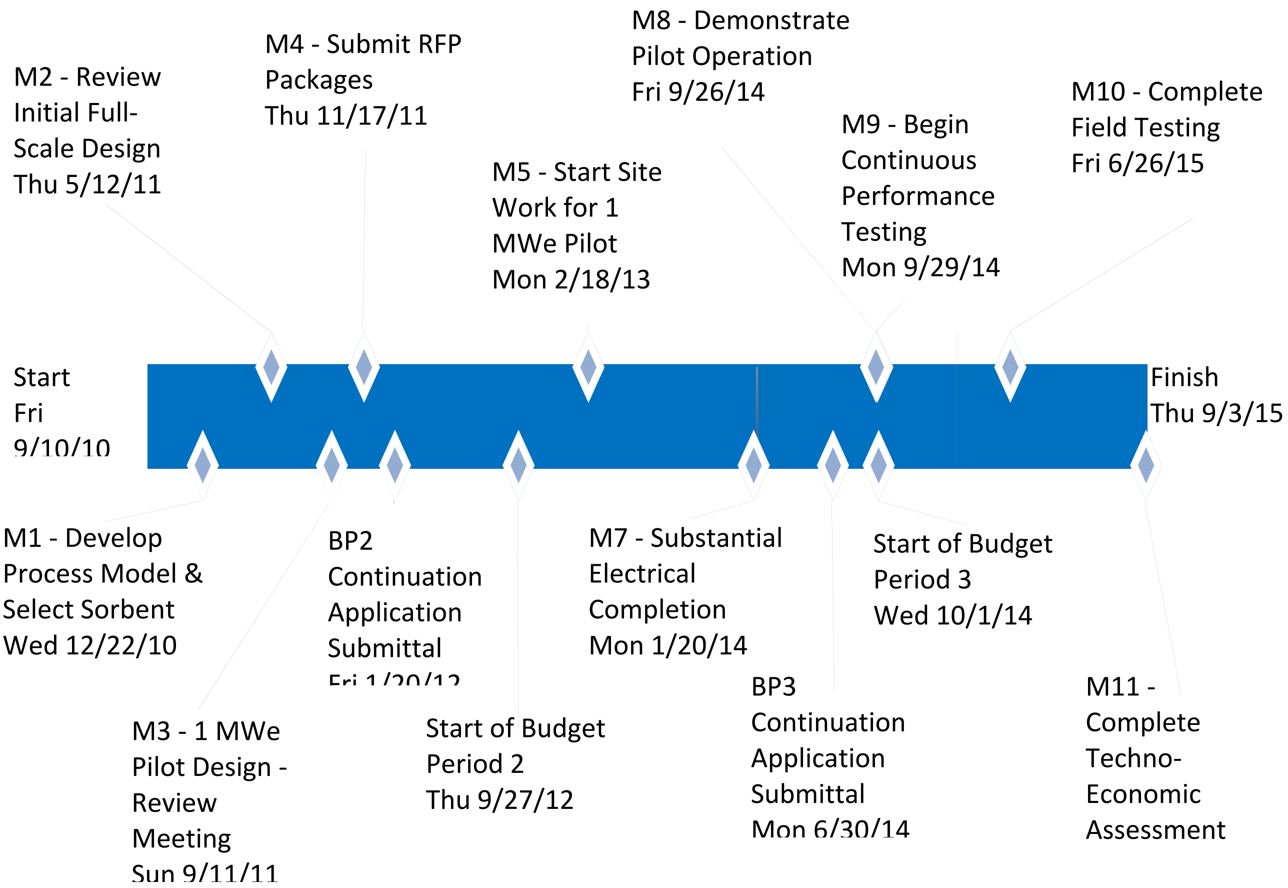


Figure 2: Project Milestone Schedule.

1.2 Objectives of the Project

The overall project objective was to demonstrate the process of solid sorbent CO_2 capture at a pilot scale using real coal-fired power plant flue gas. Specific objectives for the project included:

1. Demonstrate the technical, economic, and energy benefits of a promising CO_2 capture technology.
2. Develop performance data through the operation of a 1 MW_e pilot system on flue gas, including thermal management of sorbent and optimized system heat recovery.
3. Perform parametric testing to measure process conditions, including sorbent heat of reaction, sorbent working capacity, system pressure drop, and CO_2 capture.
4. Assess sorbent performance sensitivity to flue gas constituents.
5. Operate continuously for a target of 2 months to identify operating and maintenance issues and system reliability.
6. Remove at least 90% of the incoming CO_2 from the flue gas.
7. Produce high purity CO_2 that will be capable of meeting pipeline standards.
8. Conduct a techno-economic analysis of the commercial design, as developed by the commercial EPC.

1.3 Significance of the Project

This project was very significant in that it represented the first attempt to test solid sorbents in a continuous loop process at a “large” scale in North America. The project utilized a multistage fluidized bed adsorber with a single stage regenerator, and demonstrated for the first time that a supported amine sorbent could be used to remove 90% of the CO_2 from actual coal derived flue gas in a continuous process.

The project also highlighted some of the challenges associated with solids handling and sorbent performance and provided the necessary data to better understand fluidized bed processes in the CO_2 capture context. Finally, this project elucidated the requirements for future sorbent development necessary for solid sorbent CO_2 capture technology to be cost competitive.

1.4 DOE’s Role in the Project

The U.S. Department of Energy generously supported ADA and its partners through award DE-FE0004343 by providing 72.8% of the total funding required for the project. In addition, DOE provided administrative and technical guidance. Finally, DOE facilitated development of solid sorbent post-combustion CO_2 capture which broadens the U.S. technology portfolio for low carbon and CO_2 capture energy options.

2. Project Management

Project management and planning were managed under Task 1 of the project. This task provided time for overall project management, subcontractor and vendor management, technology transfer, and preparation of financial and administrative reports. Activities performed under this task were used to provide oversight and control throughout execution of the project, and communicate project progress with appropriate project stakeholders. Efforts within this task included:

- Generation of a Project Management Plan (PMP) per NETL guidelines and update, as necessary, to accurately reflect current status of the project as it was recognized by the award. Updates occurred to the Project Management Plan when changes occurred to:
    - The project management policies and procedures;
    - The technical, cost, and/or schedule baseline for the project;
    - Scope, methods, or approaches;
    - As otherwise required to ensure that the plan was the appropriate governing document for the work required to accomplish the project objectives.
- Planning the tests with DOE/NETL, contributing team members, and the host site.
    - Finalize host site and negotiate host site agreement.
    - Meet with plant personnel at host site and corporate and environmental personnel from the host utilities to communicate plans and coordinate all required permitting.
    - Develop a detailed test plan.
    - Develop quality assurance criteria plan and identify critical processes and metrics.
- Periodic meetings with industry representatives, cost-share participants, and DOE personnel. Meetings were scheduled to discuss progress, obtain overall direction of the program from the DOE project manager, and conduct technology transfer functions.
- Oversight and control throughout execution of the project
    - Manage and direct the project in accordance with the Project Management Plan to meet all technical, schedule, and budget objectives and requirements.
    - Manage, coordinate and report on the technical scope, budget, and schedule basis consistent with a task-oriented work breakdown structure.
    - Ensure that project plans, results, and decisions are appropriately documented to satisfy project reporting and briefing requirements.
    - Update the Project Management Plan as necessary
- Manage project risks in accordance with the risk management methodology delineated in the Project Management Plan. This included identifying, assessing, monitoring and mitigating technical uncertainties and schedule, budgetary and environmental risks. The results and status of the risk management process were presented during project reviews and in Progress Reports with emphasis placed on the medium- and high-risk items.

2.1 Project Management Deliverables

A number of reports were prepared throughout the project and submitted in accordance with the “Federal Assistance Reporting Checklist”. In addition to reports other deliverables such as technical presentations and briefings were prepared, including those to support the briefings and technical presentations indicated in Table 2 below.

Table 2. List of Project Management Deliverables.

Deliverable	Date
Project Management Plan	
Revision 1 - Initial Submittal to DOE	10/29/2010
Revision 2 - DOE Revisions	11/29/2010
Revision 3 - Revisions to Technip Scope of Work	4/4/2011
Revision 4 - Updated Resource Loaded Schedule	5/6/2011
Revision 5 - Updated Technip Scope of Work	6/6/2011
Revision 6 - Updated for BP2 CA	1/17/2012
Revision 7 - New Budget Forecast Justification	8/19/2013
Revision 7B – Included Technip CNs	8/26/2013
Revision 8 - Update for BP3 CA	6/30/2014
Revision 9 - Update for BP3 Change of Scope	3/16/2015
Monthly Progress Reports	Monthly
Quarterly ARRA Reports	Quarterly
Topical Reports	
Utility Techno Economic Assessment	3/28/2011
Industrial Techno Economic Assessment	4/15/2011
Sorbent Analysis and Selection	12/7/2011
Full Scale Costs and Energy Requirements	12/13/2011
As Built Pilot Capabilities	6/27/2014
Host Site Agreement	11/1/2013
Environmental and Construction Permits	None req'd
Preliminary Test Plan	9/27/2013
Final Test Plan	9/29/2014
Continuation Application including all supporting documentation for initiating BP2 and BP3	
Budget Period 2	1/20/2012
Budget Period 3	6/30/2014
Commercial Design and Economics Report	12/28/2014
Resource Loaded Schedule Updates	Quarterly with quarterly report

ADA provided electronic updates to the Resource Loaded Schedule on a quarterly basis to the DOE Project Officer.

Each month, a status report was prepared by ADA for the project and submitted to DOE. The report focused on the project cost and schedule status, as well as an update on project risk. Project costing and schedule performance was evaluated using earned value data and reported each month. Project performance metrics included the following.

- AC (Actual Costs Incurred)

- Cost Performance Index (CPI)
- Schedule Performance Index (SPI)
- Project Complete (based on labor projections)
- Estimated Cost to Complete (ETC)
- Variance Analysis

2.1.1 Briefings/Technical Presentations

ADA prepared detailed briefings for presentation to the NETL Project Manager. Briefings explained the plans, progress, and results of the technical effort. ADA made presentations to the NETL Project Officer/Manager at a project kick-off meeting and annual briefings.

ADA provided and presented a technical presentation and or paper at the DOE/NETL Annual Contractor's Review Meeting and at least one other technical conference each year, as approved by the NETL Project Manager.

2.2 Risk Management

There are several risks that were considered with the overall process of using solid sorbents for post-combustion CO_2 capture. The risks include the long-term stability of sorbents when exposed to flue gas, the availability of sorbents (including the number of potential suppliers), and sorbent manufacturing scale-up without impacting sorbent performance, process design to control sorbent temperatures and counteract the heat of reaction, and potential erosion and/or corrosion of process equipment. There are risks associated with CO_2 sequestration, but are considered outside the scope of the current project.

A detailed installation and start-up plan was developed between ADA and subcontractors to identify as many risks and critical path items as possible during various field activities. Objectives for this document were to quickly identify which items and tasks were on the critical path so that cost and schedule overruns could be minimized during field activities, and to set benchmark operational goals to determine when the system was ready for continual operation.

During the risk identification process, several high-level risks were acknowledged and discussed. A risk register was prepared for the project and was updated throughout the project lifecycle. High level risks for this project are provided in the appendix.

2.3 Project Success Criteria and Decision Points

The scope of work for the proposed 60-month effort covered three budget periods which corresponded to three distinct phases. Each phase has specific decision points. The decision points are described below and summarized in Table 3.

Phase I (Design Phase) = Budget Period 1 (10/1/2010 – 6/31/2012)

Phase II (Fabrication & Installation) = Budget Period 2 (7/1/2012 – 9/30/2014)

Phase III (Testing & Analysis) = Budget Period 3 (10/1/2014 – 9/30/2015)

2.3.1 Decision Points BP1 to BP2

- Full-scale design economics support technology advancement towards reducing cost of electricity increases compared to other CCS technologies. This was determined after full-scale conceptual design was complete and preliminary cost projections were developed.
- Pilot design was completed and is an appropriate scaled version of full-scale concept design.

2.3.2 Decision Point BP2 to BP3

- Pilot construction is complete
- System has been installed at field site.
- Pilot operation has been demonstrated.

Table 3. Project Decision Points and Success Criteria.

Decision Point	Date	Success Criteria
Proceed with Pilot Design Engineering	5/23/2011	Full-scale design economics support technology advancement towards reducing COE compared to other CCS technologies. This was determined after completion of preliminary cost projections developed in Task 2.
Go/No-go Decision Point (BP1 to BP2)	1/20/2012	Pilot design is complete and all necessary cost were gathered and submitted to DOE. A continuation application was submitted to DOE showing information on pilot design and cost projections for subsequent Budget Periods 2 & 3. Design was capable of removing 90% CO_2 and showed advancement towards meeting DOE's objectives as outlined in the EPEC goals.
Go/No-go Decision Point (BP2 to BP3)	10/01/2014	A continuation application was submitted to DOE prior to start of BP3. Cost and schedule information for the subsequent Budget Period was included. A detailed test plan was developed and submitted to DOE for review.
Begin commissioning and startup activities at the host site	9/1/2014	When to begin startup activities was a decision point in BP2. Criteria used to determine start date were completion of pilot installation (mechanical and electrical), loop checks and wiring checks are complete, and all interconnecting piping and auxiliary equipment has been installed. The start-up was delayed due to unavailability of flue gas from Plant Miller and water temperature from 1/1/2014 thru 9/01/2014.)
Completion of field testing during BP3	6/26/2015	The project team determined when to terminate field testing during BP3. Success criteria were met when continuous operation for at least 1 month had been completed and sufficient field testing information had been collected which were used to prepare a techno-economic assessment for the technology. The goal was to remove at least 90% of the incoming CO_2 from flue gas stream.
Project Completion	9/30/2015	Project was complete upon issuance of Project Final Report and completion of techno-economic assessment of technology. A goal of the pilot facility was to assess sorbent performance sensitivity to key flue gas constituents.

2.3.3 Project Success

- Demonstrate the technical, economic, and energy benefits of a promising CO_2 capture technology.
- Develop performance data through the operation of the system on flue gas, including thermal management of sorbent
- Assess sorbent performance sensitivity to key flue gas constituents
- Operate continuously for a target of two months to identify operating and maintenance issues and system reliability
- Remove at least 90% of the incoming CO_2 from the flue gas
- Produce high purity CO_2
- Conduct a techno-economic analysis of the commercial design

3. Preliminary Design Activities: Process Technology and Sorbent Selection

The fundamental process design and final sorbent selection were conducted during Task 2 of the project through the preliminary design activities outlined by the subtasks listed below.

- Subtask 2.1: Review Sorbent Characteristics, Field Test Data, and Viability Design Assessment
- Subtask 2.2: Process Calculations and Modeling

An optimized post-combustion process is strongly dependent on both process and sorbent characteristics. Thus, these activities were considered simultaneously during the preliminary design activities. This was an iterative effort because the sorbent characteristics strongly influence the process design details.

The following sections provide a summary of the process design considerations and sorbent evaluation efforts, followed by a description of the conceptual process design.

3.1 Process and Sorbent Selection

3.1.1 Reactor Design: Viability Assessment

There are several key considerations when selecting a reactor design, including:

- Capital cost
- Footprint
- Pressure drop
- Gas solids contacting
- Operability (i.e., maintenance frequency and annual cost)
- Constructability

It is worth noting that effective heat transfer and effective gas solids mixing are not decoupled; a well-mixed system should accomplish both if designed correctly. Reactor types considered during the conceptual design effort included:

- Fixed beds – stationary systems with internal heat transfer
- Entrained reactors – sorbent simultaneously reacts and is conveyed using the flue gas in the adsorber and (most likely) by a mixture of CO_2 and H_2O in the regenerator
- Moving beds – densely packed sorbent moves in the opposite direction as the gas while the gas can flow either countercurrent or across the sorbent
- Staged fluidized beds – a series of fluidized beds in the bubbling regime where the gas moves upwards while the sorbent enters at the top bed and leaves the adsorber in the bottom bed
- Trickle down reactor – sorbent with significant mobility (i.e., not packed) flows counter current to gas

Some reactors were quickly and easily removed from consideration. For example, an entrained reactor cannot be effectively used to attain acceptable CO_2 loadings. Because the gas entrains the sorbent and 90% CO_2 capture is required, the sorbent will, at best, reach an equilibrium loading dictated by a CO_2 partial pressure of 0.015 bar. Assuming an adsorption temperature of 40°C and a regeneration temperature of 120°C and a $PCO_2 = 0.81$ bar, the working capacity in an entrained reactor would be approximately 1.5 g CO_2/100 g fresh sorbent, which is significantly lower than the reported working capacity of aqueous MEA.[20]

True counter-current flow, such as a trickle-down reactor where solids are introduced at the top of a column and gas is introduced at the bottom, was considered but was eliminated as an option when considering the complexities associated with scale-up to a full-scale power plant. Novel packing structures would be required to maintain good sorbent distribution and effective gas/solids mixing. In the midst of the packing for distribution, heat transfer surface area would be required, which would further crowd the space in the reactor. Optimally the system would become a fluidized bed or moving bed system, which were already under consideration, so the trickle-down reactor concept was abandoned.

Significant heat transfer surface area is required to operate a TSA-based CO_2 capture process, but the exact amount of heat transfer surface area depends on the reactor type. Widely known empirical correlations were utilized to estimate the heat transfer coefficients for moving beds and fluidized beds. Because this is an order-of-magnitude estimate, it is assumed that it applies to both fixed beds and tightly packed moving beds because the mechanisms of heat transfer are largely similar.

For fixed beds, Li and Finlayson[21] provided the following empirical correlation:

$$\frac{\mathbf{h_w D_p}}{\mathbf{k_p}} = \mathbf{0.17 Re_p}^{\mathbf{0.79}} \qquad (1)$$

Where h_w is the heat transfer coefficient, D_p is the particle diameter, and k_p is the thermal conductivity of the particle. Equation 1 is applicable as long as the particle Reynolds number (Re_p) is in the range of $20 < Re_p < 7600$, where:

$$\mathrm{Re_p} = \frac{\rho_f v D_p}{\mu_f} \tag{2}$$

The properties of the flue gas are known from the composition, temperature, and pressure while the properties of the sorbent can be estimated for illustrative purposes only and are collected from several sources.[22-25] These values are provided in Table 4. Different particle sizes were used for the fixed/moving bed analysis and fluidized bed analysis. For the purpose of these calculations, the sorbent properties remain constant with the exception of particle size. For the sorbent particle size it is unreasonable to propose a fixed bed and a fluidized bed with the same particle size, so the particle size was reduced by a factor of ten for the fluidized bed.

Table 4. Flue Gas and Sorbent Characteristics for Sorbent Used as Basis for 1 MWe Pilot.

Property	Units	Value
ρ_f	kg/m^3	1.04
v	m/s	1.2
μ_f	Pa·s	1.9e-5
k_f	W/m-K	0.025
D_{AB}	m^2/s	2.6e-5
D_p-fixed	m	0.001
D_p-fluidized	m	0.0001
k_p	W/m·K	0.08

The $\mathrm{Re_p}$ can be calculated using Equation (2) and the values in Table 4. Under the proposed conditions, $\mathrm{Re_p}$ is approximately 67, so the correlation shown in Equation (1) is valid. Solving for the heat transfer coefficient, h_w, in Equation (1) yields a value of 350 W/m^2·K, which can be used as an order-of-magnitude estimate for the fixed bed and moving-bed systems under evaluation.

There are also many correlations available for estimating the overall heat transfer coefficient in fluidized beds. Often, such correlations or data compilations link the Nusselt number to the $\mathrm{Re_p}$ defined in Equation (2). The Nusselt number of interest can be defined by:

$$\mathrm{Nu} = \frac{h_c D_p}{k_f} \tag{3}$$

Again the $\mathrm{Re_p}$ can be calculated using Equation (2). Everything in the calculation is the same except the particle diameter, which has been assumed to be an order of magnitude smaller than that which would be used in a fixed bed (i.e., 100 µm versus 1 mm). An integrated plot of several heat transfer correlations compiled by Zenz and Othemer[26-28] can be used to estimate Nu based on $\mathrm{Re_p}$. Assuming a fluidized bed in the bubbling fluidized bed regime with a void space of approximately 0.7, the Nu is estimated to be three, which results in an overall heat transfer coefficient, h_c, of 743 W/m^2·K.

Based on the calculations completed, the heat transfer coefficients of fixed and moving beds can be projected to be only half that of a fluidized bed. To understand whether such a large difference in the ratio of heat transfer coefficients is expected, experimental work for a different system can also be reviewed. Figure 3 was collected by Xavier et al.,[29,30] to measure the effect of pressure on heat transfer between a flat surface and glass spheres in an N_2 atmosphere. Although this system differed significantly from the conditions that would be present during post-combustion CO_2 capture, the ratio of the overall heat transfer coefficient for fixed/moving beds and fluidized beds can be observed from the data. As the gas velocity increases, the heat transfer coefficient in the fixed bed slowly increases. Then, as the bed achieves the minimum fluidization velocity (in the lowest pressure case this occurs at about 0.23 m/s) the overall heat transfer coefficient increases as a step change from approximately 30 to 270 $W/m^2 \cdot K$. This large step change clearly demonstrates the advantage of using a fluidized bed for CO_2 capture that will require significant heating and cooling; based on either the data shown in Figure 3 or the correlations used above, the heat transfer surface area for a fluidized bed would be significantly less than that required for a moving bed or fixed bed.

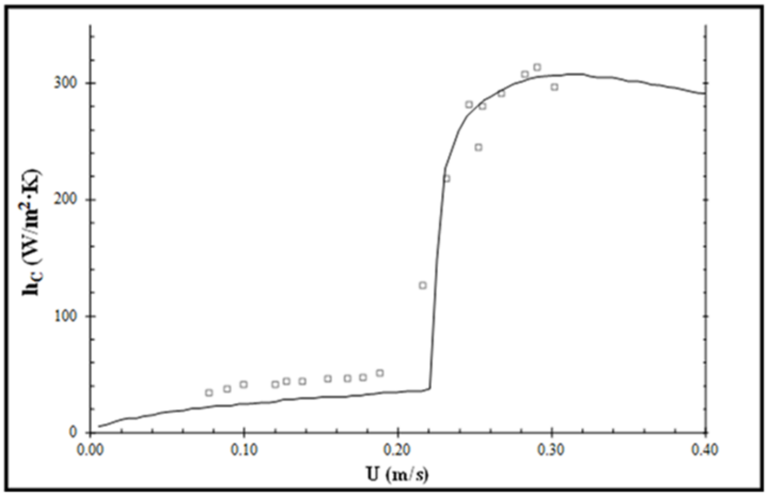


Figure 3. Experimentally Measured Overall Heat Transfer Coefficient. The lowest pressures (open squares) are primarily of interest. Specifically the purpose of this figure is to illustrate the step change in the overall heat transfer coefficient when a material is fluidized. Xavier et al.,[29,30]

The overall objective for the design of a CO_2 capture process is not maximizing the heat transfer coefficient, but rather to minimize the overall increase in the levelized cost of electricity. The overall increase in the cost of electricity can be attributed to two factors: 1) capital costs and 2) operating costs (primarily due to the energy penalty).[20] It was revealed within this project's initial TEA that the installed capital costs and the energy penalty were similar in magnitude. Therefore, neither should be neglected during technology development. The operating costs were largely the same for different reactor types as long as the sorbent loading was maximized, excluding the impact of pressure drop on operating costs. To assess the capital costs, order-of-magnitude quotes from vendors were obtained for moving and fluidized beds using the supported amine sorbent. It quickly became clear that the amount of required heat transfer surface area had a dramatic impact on the capital costs. For staged fluidized beds, the total adsorber height as well as the height of each stage was dependent on the heat transfer surface area. In the initial TEA, the height of the vessel was primarily determined by the heat transfer surface area rather than by the kinetics of the primary amine functionalized adsorbent. This is a typical design outcome for fluidized bed reactors used for decades in the chemical, petrochemical and energy industries.

Regarding overall height, the moving beds did exhibit a few advantages. There was no need for headspace as was the case above each bed in the fluidized reactor. Rather, the sorbent would be packed relatively densely as it moved past the heat transfer surfaces. However, due to the lower heat transfer coefficients, the heat transfer surface area required was significantly greater than that of the staged fluidized beds, so the total number of reactors was significantly greater, as were the capital costs.

It is well known that the required heat transfer surface area has an impact on reactor cost. In fact, Peters and Timmerhaus[31] recommend using correlations for heat exchangers to estimate relative costs for reactors that are also responsible for large amounts of heat transfer. The data extracted for Figure 4 is from Peters and Timmerhaus[31] and shows the relative cost for heat exchangers. Note that the surface area and costs are plotted on a logarithmic axis; the costs increase significantly as the heat exchange surface increases. ADA and project partners actually obtained vendor quotes when comparing moving bed and fluidized bed heat exchangers. While these quotes are considered confidential and cannot be included in this report, they support the conclusion that the significantly lower heat transfer surface area required for fluidized beds results in lower overall capital costs.

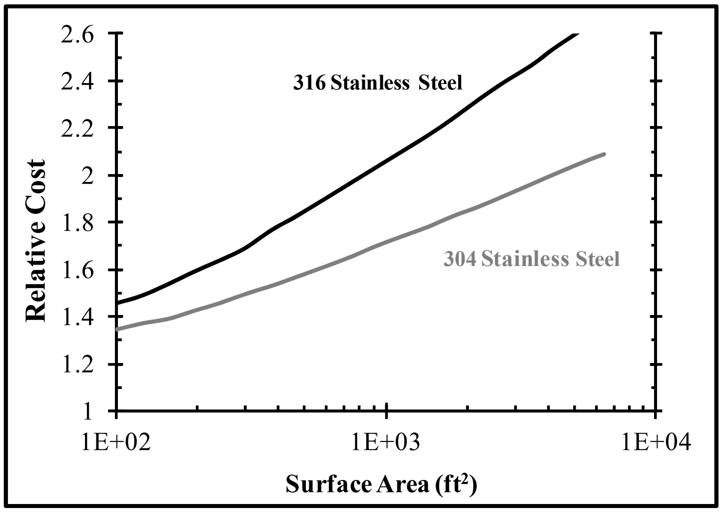


Figure 4. Relative cost for heat exchanger based on surface area from Peters and Timmerhaus.[31]

It has clearly been shown that the overall heat transfer coefficient in a bubbling fluidized bed is superior to that of other reactor types and that translates into lower surface area. For this reason, a bubbling fluidized bed was selected as the optimal reactor type. However, a single fluidized bed is not optimal for adsorption. Rather, multiple stages of fluidized beds must be used to approach a counter current flow and increase the loading on the sorbent.

3.1.2 Sorbent Selection and Characterization

Selecting the optimal sorbent characteristics for successful implementation across the coal-fired power sector requires consideration of several factors. Furthermore, it is not a simple task to quantitatively compare different adsorbents using laboratory tests in a manner that can be extrapolated to inform long-term, full-scale decisions. Criteria for comparison can be developed, based on fundamental considerations.

More than 250 sorbents were screened in ADA's laboratory, either in a fixed bed or in a thermogravimetric analyzer (TGA), prior to initiation of this project and this data was used to identify appropriate candidate sorbents for scale-up in a TSA process. Limited sorbent screening followed by extensive laboratory characterization was completed during this program to select a sorbent type appropriate that would be optimal in a TSA process, and to characterize that sorbent in the detail required to provide required information to estimate full scale economics and design specifications for the 1 MWe pilot system. The key sorbent characteristics identified for process design consideration are listed below.

- CO_2 working capacity
- Cyclic stability
- Reaction kinetics
- Effect of flue gas constituents including SO_2 and moisture
- Heat of Reaction
- Resistance to attrition
- Physical characteristics: particle size distribution and density
- Sorbent cost
- Sorbent fluidization and handling properties
- Heat transfer coefficient

At the onset of this project, two types of sorbents were under consideration: activated carbons and supported amine sorbents. Supported amine sorbents were ultimately selected because they were able to capture more CO_2 by weight in a temperature-swing process between adsorption and regeneration conditions. Using this type of sorbent in a fluidized-bed process resulted in lower sorbent circulation rates in a fluidized-bed process design and expected lower total regeneration heat duty. If activated carbon were selected, a different process design, such as a fixed monolithic design operating in a TSA environment with a slight pressure swing, may have provided advantages.

Once the sorbent type was selected, a specific sorbent was identified and characterized so that design specifications for the 1 MWe pilot could be finalized. Three promising supported amine sorbents were evaluated and compared. An ion exchange resin functionalized with a primary amine was selected due to its relatively high CO_2 capacity in a temperature-swing process, acceptable particle size, stability in the presence of moisture, stability at temperatures of 120°C or greater, and the ability to purchase and/or produce the material on the scale required for the 1 MWe pilot.

After the specific sorbent was selected, it was characterized using various experimental equipment and techniques including thermogravimetric analysis (TGA), mass spectrometry (MS), differential scanning calorimetry (DSC), and jet cup attrition tests.

3.2 Key Sorbent Selection Criteria

There are several key selection criteria that were used to compare different sorbents.

3.2.1 CO_2 Delta Loading

The CO_2 delta loading (or the working capacity) is considered the most important selection criteria for sorbents and is directly related to how much sorbent must be circulated, which affects the regeneration energy (sensible heat), the conveying requirements, and the sizes of the adsorber and regenerator vessels. The CO_2 delta loading (g CO_2/100 g fresh sorbent) is defined as follows:

$$CO_2\ Delta\ Loading = 100 * \frac{m_{CO2-adsorption} - m_{CO2-regeneration}}{m_s} \tag{4}$$

where;

m_{CO2}-adsorption is the CO_2 loading at the capture/adsorption conditions, g

m_{CO2}-regeneration is the CO_2 loading at regeneration conditions, g

m_s is the sorbent mass, g

Note that the terms delta loading and working capacity are often misused in the public literature. To determine a CO_2 delta loading, the adsorption and regeneration temperature and CO_2 partial pressure must be specified. For the purposes of the evaluations discussed in this report, the adsorption conditions are characterized by low temperature (40 to 60°C) and low CO_2 partial pressure (0.1 to 0.2 bar). The regeneration conditions are characterized by higher temperatures (greater than 90°C) and higher CO_2 partial pressure (0.8 to 1.0 bar). For the purposes of comparing different sorbents, the CO_2 delta loading is calculated using the equilibrium conditions at these temperatures and partial pressures. However, in an actual commercial system the sorbent will likely not achieve equilibrium loading due to factors such as limited residence time, slight variations in temperature, and varying partial pressure of CO_2 in a full-scale system. A qualitative assessment of laboratory results that can indicate changes in reaction time as a result of changes in temperature should be incorporated into a commercial design process. Any commercial system will then need to be optimized for cost and performance.

3.2.2 Cyclic Stability

The act of repeatedly heating and cooling some sorbents leads to the slow degradation of the material. For some supported amines, volatilization of the amine can occur due to only a temperature swing, which is considered unacceptable for a commercial process. Cyclic stability can be measured in the laboratory, but requires a significant time commitment and can therefore be easily overlooked. Under

DE-FE0004343 cyclic stability was considered a key selection criterion and was assessed at realistic regeneration conditions.

3.2.3 Reaction Kinetics

One important reason that a sorbent may not achieve equilibrium CO_2 loading in an operating system is due to kinetics. Unfortunately, measuring kinetics is not always straightforward. The rate of CO_2 uptake can be highly dependent on temperature, CO_2 partial pressure, sorbent particle size, gas/solids contacting, etc. For this reason, the sorbent kinetics must be determined under conditions that are relevant to the final process under consideration. If assessed properly, the kinetics can be used to determine many important process factors, such as required contacting time, equipment size, contacting scheme, etc.

Kinetic information was obtained from TGA data and from trickle down reactor data. The rate that a sorbent will remove CO_2 in a fluidized bed is governed by the mass transfer of the CO_2 in the gas to the sorbent, which is fairly efficient in a fluidized bed, the rate that the sorbent reaches the adsorption temperature, and the reaction kinetics of the sorbent.

The first test to assess kinetics was using the trickle-down reactor, a 2.29 m (7.5 ft) long, heated drop-tube reactor. Sorbent, at the adsorption temperature, was introduced at the top of the tube and a mix of nitrogen and CO_2 at the desired concentration was introduced at the bottom. The CO_2 concentration was measured at the top of the tube where the gas was exiting. The sorbent residence time in the tube was less than 2 seconds. Unlike the TGA tests where the temperature was changed to initiate a new test condition, for the drop tube tests, the sorbent was at the test temperature and both the CO_2 concentration and the flow conditions were changed. Although the reaction between the sorbent and CO_2 was exothermic in both cases, the base material was at the correct adsorption temperature at the beginning of the drop tube test.

The TGA is designed to measure equilibrium conditions, but some indications of reaction rate can be extrapolated from the rate of mass change over time as conditions are modified. In understanding how to interpret kinetic data, it is important to understand how long it takes sorbent particles to equilibrate with the gas. For a spherical particle with radius R_p, the characteristic time for relaxation[32] of the temperature profiles in the gas phase around the particle is

$$\tau_h = \frac{{R_p}^2}{\alpha} \tag{5}$$

where $\alpha = {k_f}/{\rho_f C_p}$ and k_f is the thermal conductivity of the gas, ρ_f is the gas density, and C_p is the gas heat capacity. For characteristic times approaching the relaxation time or shorter than it, the system cannot follow the imposed temporal changes. Similar to heat transfer, the characteristic time for relaxation of concentration profiles in the gas phase around the particle is

$$\tau_m = \frac{{R_p}^2}{D_{AB}} \tag{6}$$

where D_{AB} is the diffusivity of species A in a gas mixture of A and B. The characteristic time for heat transfer within the spherical particle is

$$\tau_{hp} = \frac{{R_p}^2}{\alpha_p} \tag{7}$$

Where $\alpha_p = {k_p}/{\rho_p C_{pp}}$ and k_p is the thermal conductivity of the particle, ρ_p is the particle density, and C_{pp} is the particle heat capacity. For a 100 micron spherical particle of Sorbent BN in a gas of CO_2 and N_2 at 120°C, the characteristic times are:

$\tau_h = 1.57 \times 10^{-6}$ seconds

$\tau_m = 3.85 \times 10^{-6}$ seconds

$\tau_{hm} = 0.02$ seconds

The characteristic time for equilibration of the temperature *inside* the particle is much greater than the characteristic time for equilibration of temperature or concentrations profiles around the particle. Thus, if there is a release of heat on the particle surface, the heat will be largely conducted out into the gas.

The time scales of adsorption or desorption in the TGA are on the order of minutes. Therefore, the temperature BN particles in the TGA will have sufficient time to equilibrate as temperature is changed in the TGA.

In order to utilize TGA results, it is valuable to estimate the mass transfer coefficient and the kinetic rate coefficient to determine whether the process is kinetically or mass transfer limited in the test apparatus. As discussed by Fogler in *Elements of Chemical Reaction Engineering*, mass transfer uses similar relationships to heat transfer.[33] One of the most important relations is the Frossling correlation:

$$\boldsymbol{Sh = 2 + 0.6Re^{\frac{1}{2}}Sc^{\frac{1}{3}}} \tag{8}$$

Equation 4 yields the Sherwood number, which is dimensionless and is a function of the dimensionless Reynolds number and Schmidt number. The Sherwood number for a spherical particle may be expressed by:

$$Sh = \frac{K_C D_p}{D_{AB}} \tag{9}$$

in which K_c is the mass transfer coefficient, D_p is the particle diameter, and D_{AB} is the diffusivity. The Schmidt number is given by:

$$Sc = v/D_{AB} \tag{10}$$

whereby ν is the kinematic viscosity of the bulk gas and D_{AB} is the diffusivity of the gas.

Rearranging these equations it is possible to solve for K_c. Even at a lower limit of the Reynolds number equal to 0, which is the worst case for mass transfer, the value of K_c approaches a value of 0.5 m/s for a particle size of 100 microns, which is well within the pilot sorbent size distribution. The number is not as important as the order of magnitude, which will now be explained.

Fogler also provides a relation for a simple model of one reacting gas species on a spherical particle if the particle is kinetically limited:

$$-r_{As}" = \frac{K_r C_A}{(1+K_r/K_C)} \approx K_r C_a \qquad (11)$$

$-r_{As}$" is the rate of disappearance of species A per unit area of particle surface in the units of mol A/m_2*s.[33] The value of $-r_{As}$" may be calculated using TGA data whereby the number of moles of CO_2, in this case species A of interest, has adsorbed on sorbent particles' surfaces. Using the known sorbent density, an estimate of 100 micron size particles, and the mass of sorbent loaded in the TGA, it is possible to estimate a sorbent surface area. Since the duration of the TGA test is also known, it is possible to calculate the moles of CO_2 adsorbed per square meter each second to determine the value of $-r_{As}$". Since the concentration of CO_2 is also known in the TGA, the value of C_a is also known, or in this case, C_{CO2}.

Using this information, it is possible to solve for K_r. In this particular example, $K_r = 10^{-6}$ m/s, which is five orders of magnitude smaller than the corresponding value for K_c in an artificial worst-case scenario with no flow, indicating that the sorbent in the TGA is clearly not mass-transfer limited and equation 11 is appropriate and valid.

It is important to note that in the TGA there is flow of gas past the pan of sorbent, promoting mass transfer of CO_2 from the bulk gas to the area around the pan. The models presented above indicate that mass transfer is not a limiting factor in any practical analysis of this sorbent, which is why working capacity of sorbent measured in the TGA was nearly equivalent regardless of whether the sorbent had been milled, was beaded into large 1 mm particles, had been used in the pilot, or was a manufactured sample investigated in the lab.

3.2.4 Effect of Flue Gas Constituents including SO_2 and Moisture

Through previous evaluations, it was determined that all sorbents that chemically react with CO_2 will also chemically react with SO_2. In the case of supported amine sorbents, the chemical bond between the SO_2 and the amine is considered a heat-stable salt because it will not decompose in a simple temperature swing. This is also a concern for aqueous amine capture processes. The effect of flue gas constituents must be considered when selecting sorbents, because the cost implications of replacing and/or treating sorbents for reclamation could be considerable.

During previous sorbent screening, it was determined that some sorbents, such as zeolites, preferentially adsorb moisture over CO_2. Such sorbents were not considered or evaluated under DE-FE0004343. Only materials that could effectively remove CO_2 in the presence of saturated conditions were considered feasible options. However, even such sorbents may adsorb and release H_2O under the adsorption and regeneration conditions of the CO_2 capture process. Because the uptake of H_2O during adsorption and release of H_2O during regeneration contribute additional energy requirements to the CO_2 capture process, it is important to quantify the uptake of H_2O on sorbents at realistic operating conditions.

3.2.5 Heat of Reaction

The reaction between amine functionalized sorbents and CO_2 is exothermic, thus heat is generated. Likewise, when the sorbent is heated to promote the release of CO_2 during regeneration, heat input is required because the release of CO_2 is endothermic. Understanding these requirements for heat input is crucial for designing the overall CO_2 capture process. An external laboratory was hired to measure the heat generated by reacting sorbent with CO_2 using calorimetry. Because the calorimeter was not integrated with a thermogravimetric analyzer, the isotherms generated separately with the TGA were used to estimate the CO_2 uptake at the different calorimeter test conditions.

3.2.6 Resistance to Attrition

Eventually some sorbents will be physically broken down due to the mechanical wear and tear of being circulated. Depending on the type of system utilized for the capture process, the attrition may be defined differently. In general, a sorbent has been unacceptably attrited when it is undesirably entrained by process gas streams (either during adsorption or regeneration). At this point it will no longer be useful to the CO_2 capture process and must be separated from the gas stream by a baghouse or other means. The rate of attrition is related to the physical strength of the sorbent and the process in which the sorbent is being used. The physical strength of the sorbent is considered a key selection criterion because high attrition levels would result in sorbent replacement rates that would negatively impact the overall process economics. Quantifying attrition can only be accomplished by operating for long periods of time in the actual CO_2 capture system. However, there are several different types of laboratory-scale tests that can provide qualitative comparisons of the physical durability of different potential sorbents, including jet-cup attrition and crush strength.

Jet-cup attrition testing is a common method for evaluating the friability of solids, such as catalysts, and evaluating particle attrition in fluidized beds.

Crush strength provides a quantitative measurement of particle hardness. However, it does not provide a means to directly predict attrition because this is a highly process-dependent quantity. For the most promising beaded material provided to the sorbent screening program, the crush strength was measured.

3.2.7 Physical Characteristics: Particle Size Distribution and Density

Particle densities were measured for two promising sorbents. The testing was conducted using a proprietary experimental method. The particle density is defined as the mass per unit particle. This quantity is necessary for determining fluidization and entrainment related properties.

To maintain proper fluidization within the adsorber, the sorbent particles are sized to retain a Geldart A distribution for fluidized beds at up to 1.4 m/s (4.6 ft/s) gas velocity.

3.2.8 Sorbent Cost

Sorbent cost is a key selection criterion due to the importance of this term on the overall process economics. Because no large-scale market exists today for many of the sorbents being evaluated for CO_2 capture, it is difficult to get an accurate assessment of commercial sorbent costs. However, it is possible to gain order of magnitude estimates based on raw materials and processing complexity. The base sorbent cost used for economic modeling was \$5/lb , based on raw material costs and aggressively limiting scale-up manufacturing costs.

3.2.9 Sorbent Fluidization and Handling Properties

Adequate sorbent fluidization and conveying are critical to success with a fluidized bed. For example, if particles agglomerate or demonstrate other characteristics that are unpredictable or not conducive to stable fluidization or handling, operation of the overall system will be affected.

3.2.10 Heat Transfer Coefficient

The heat transfer coefficient of a sorbent is a function of both the process and the sorbent. The heat transfer coefficient measured in a fixed bed of sorbent will differ from the heat transfer characteristics in a fluidized bed. The metric of interest is the behavior in the actual full-scale process.

3.3 Experimental Methods

3.3.1 Trickle Down Reactor

The isotherms developed using the TGA are useful to determine optimal process conditions (i.e., temperatures and CO_2 concentrations). However, these laboratory tests are insufficient to fully answer questions about kinetics, required contact times etc., in different types of solid/gas contact devices. One of the reactor configurations considered for this project was a counter current (i.e., trickle down) reactor for adsorption, where the flue gas velocity would be low enough so as not to entrain the sorbent particles. ADA constructed and operated a bench-scale counter current reactor to identify the required contact time for sorbent BN to remove 90% of CO_2 from simulated flue gas. The objectives of the test included the following:

- Determine the approximate height of the adsorber required for 90% CO_2 removal
- Qualitative assessment of sorbent kinetics of adsorption

A schematic of the test apparatus is provided in Figure 5. The adsorber section consists of an insulated 5 cm ID (2 inch ID), 2.29 m (7.5 ft) long section of polycarbonate tubing with temperature and gas sample ports at the top, middle, and bottom. Simulated "flue gas" was created by blending nitrogen and CO_2 and passing the gas stream through a humidifier. The gas was then heated to the required temperature in the heating section prior to entering the sorbent collection barrel. The gas then flowed up the adsorber section where it came in contact with the downward flowing sorbent. Sorbent was introduced to the adsorber section by a vibratory feeder at a controlled rate. The sorbent passed through two (2) distribution screens prior to entering the adsorber to distribute the sorbent evenly across the adsorber cross-section. The mean diameter of the sorbent was 0.9 mm and the particle density was approximately 385 kg/m^3.

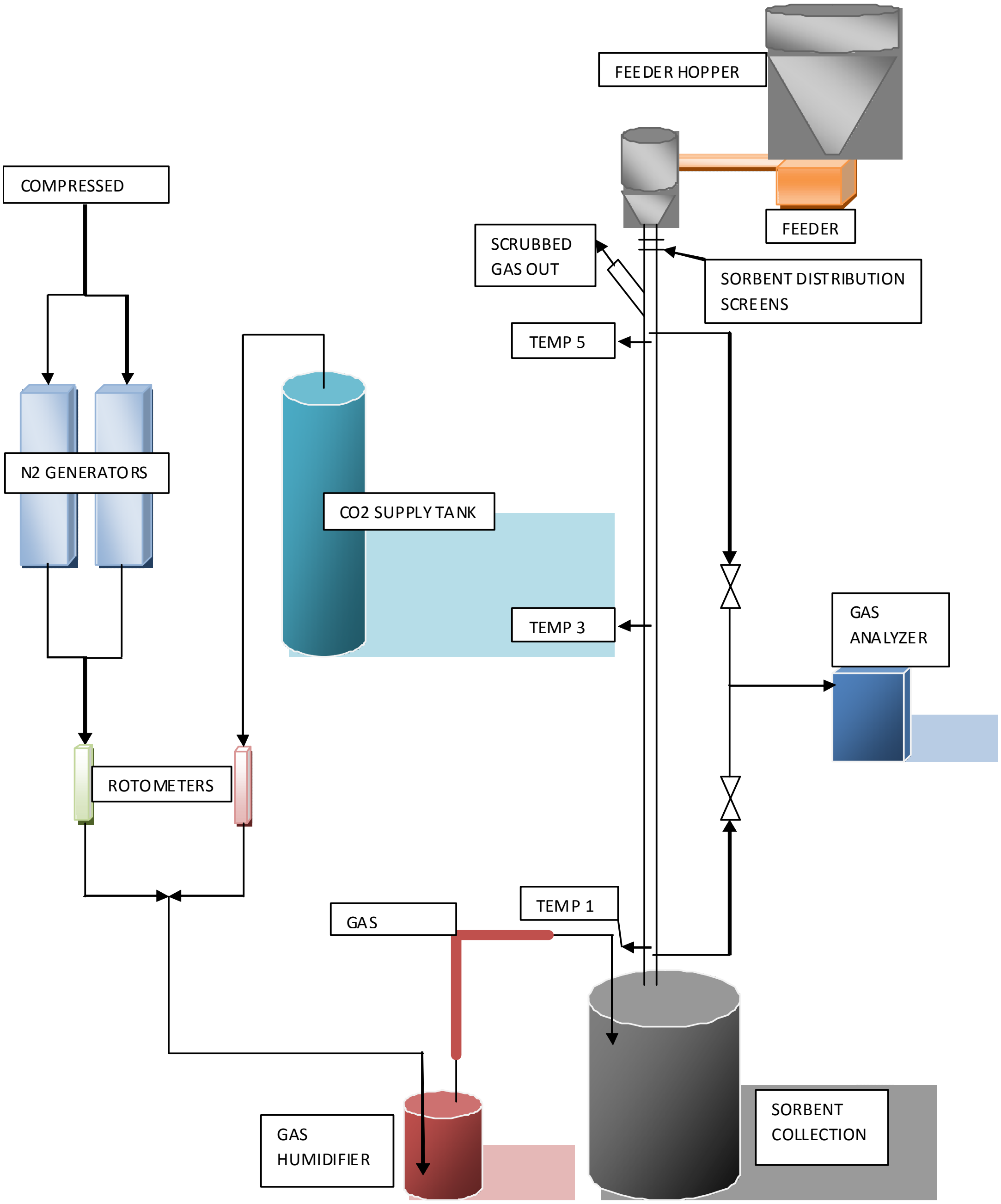


Figure 5. Counter Current Contactor (Trickle Down Reactor) Testing Schematic.

3.3.2 Thermogravimetric Analyzer

A thermogravimetric analyzer (TGA) was used to measure the CO_2 loading for select sorbents under different temperatures and CO_2 partial pressures. The TGA used during sorbent evaluation was a Perkin Elmer Pyris 1. The TGA was operated at less than standard atmospheric pressure, because the tests were conducted at ADA's headquarters in Highlands Ranch, CO (elevation 1643 m); therefore, without additional pressurization of the apparatus, when 100% CO_2 gas was used, the CO_2 partial pressure was only 0.81 bar. Note that a small amount of moisture was added to the gas while using the TGA in order to introduce a humid environment, which the sorbent would experience in practice. Unless moisture uptake was specifically being evaluated, the moisture levels were approximately less than 1% by volume even though flue gas moisture content could be up to 10% moisture. One of the main reasons for this limitation is that in a TGA there are very small diameter tubes that convey gas streams. If there is a cold spot along the tubes, then water may condense and invalidate the test or cause damage to the tubes. For this reason, humidified testing was conducted below saturation. The test details are provided in Table 5.

Table 5. Typical TGA Operating Conditions.

TGA	Sample Size (mg)	Gas Flow Rate (mL/min)	Temperatures Evaluated (°C)	CO_2 Partial Pressure (bar)	Dew Point (°C)
Perkin Elmer Pyris 1	1-5	100	40 to 170	0, 0.04, 0.081, 0.15, 0.5, 0.81	<22 to 45

A mass spectrometer (MS) was installed downstream of the TGA for certain experiments to measure the gas concentrations, including potential decomposition products of the amine, if applicable. The test setup is shown in Figure 6 with the TGA on the right hand side of the picture, the MS in the middle and the computer used for data acquisition on the left hand side of the picture.

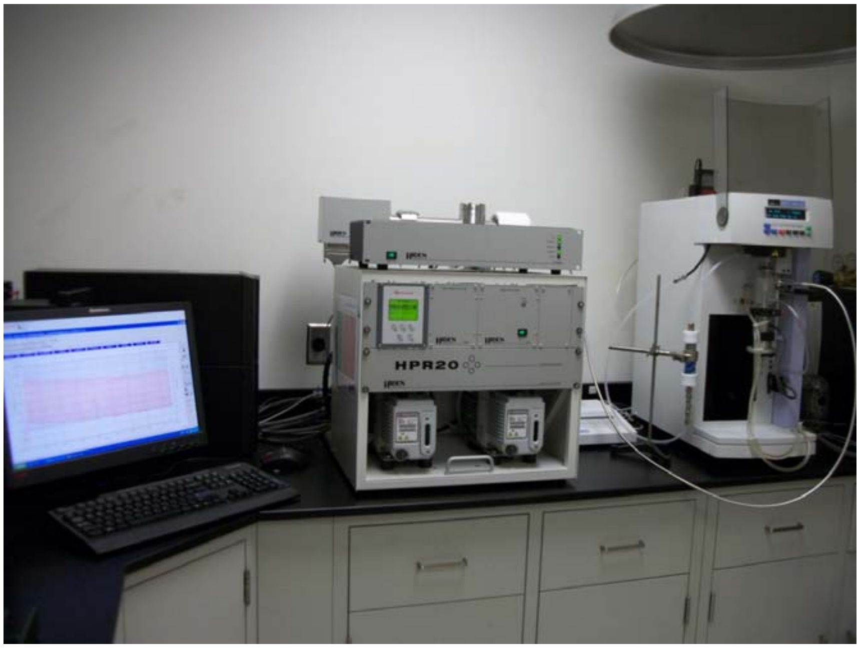


Figure 6. Laboratory Experimental Equipment: TGA (Right) and Mass Spectroscopy (Left).

3.3.3 Fixed Bed Reactors

A specialized fixed bed reactor was designed to be used in the laboratory on simulated flue gas as well as in the field on actual flue gas. The system was fully temperature controlled and incorporated a series of automated valves to allow cycling between adsorption and regeneration conditions. A Programmable Logic Controller (PLC) was employed to completely automate the testing process. A sketch of the system configured for either simulated gas in the laboratory or actual flue gas in the field is presented in Figure 7.

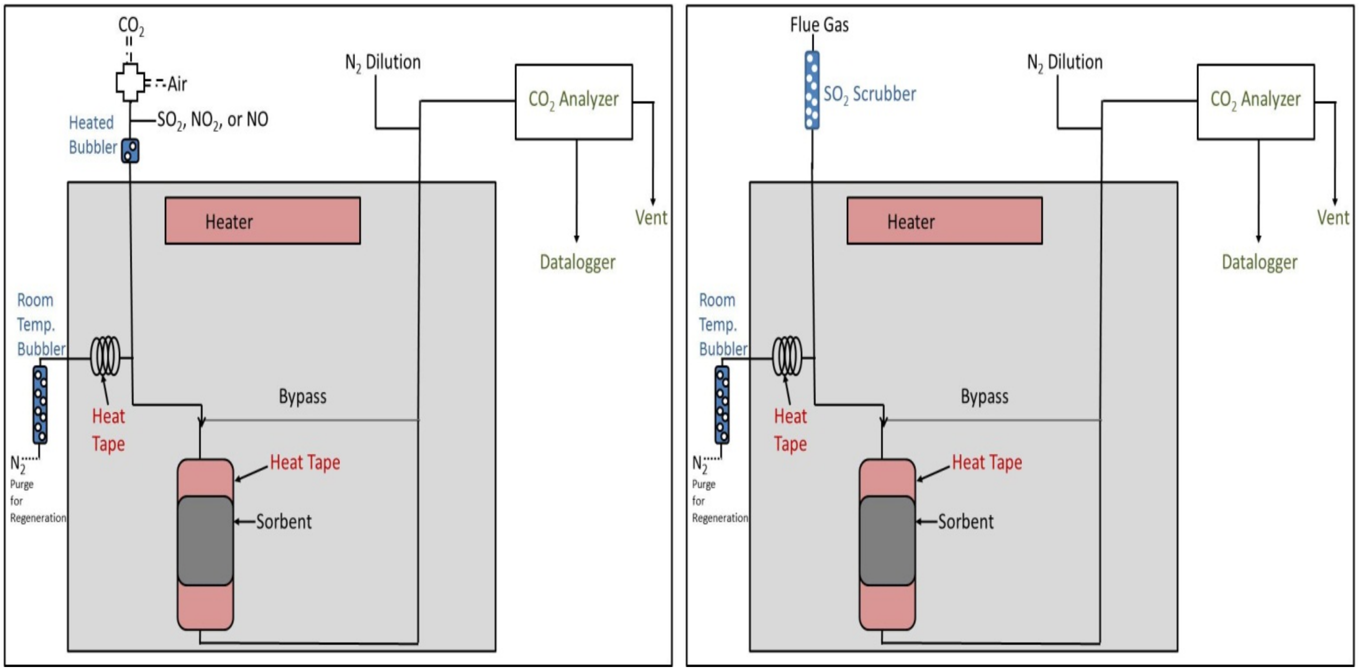


Figure 7. Sketch of the Fixed Bed Sorbent Screening Test Unit Configured for the Laboratory (Left) and Field (Right).

The flow rate of either simulated or actual flue gas was approximately 300 mL/min, and the amount of sorbent in the reactor was typically in the range of 0.4 to 2.5 g. The regeneration profile was measured while the sorbent was heated under a nitrogen (N_2) purge so that observation of the regeneration breakthrough curve could be monitored for degradation due to temperature cycling. An example of the breakthrough curve during adsorption, and the regeneration profile, are shown in Figure 8.

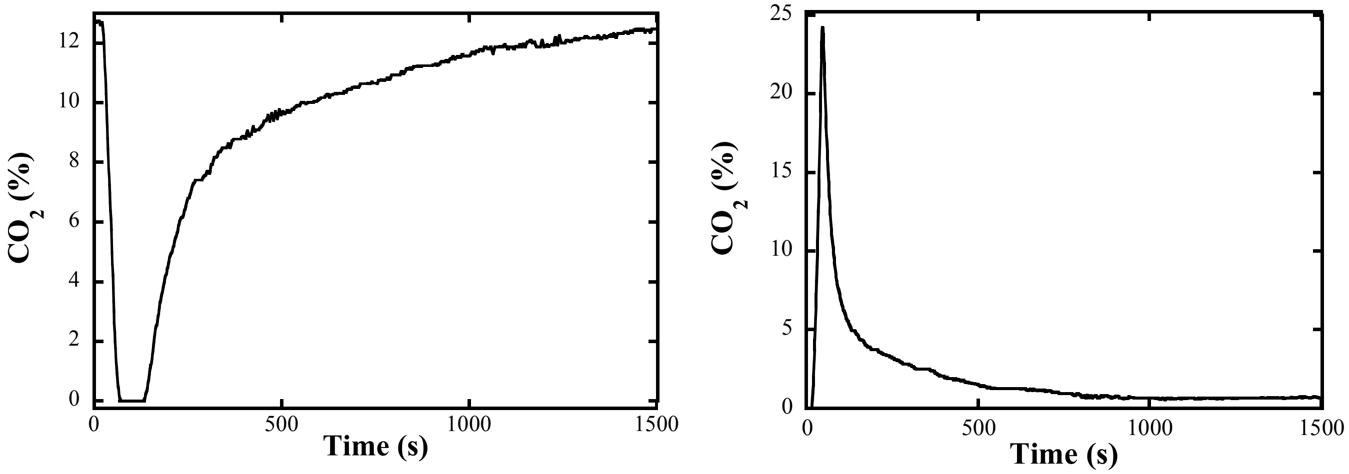


Figure 8. Example of Fixed Bed Adsorption (left) and Desorption (right) Breakthrough Profiles.

Tests conducted using the fixed-bed device are focused on measuring the cyclic stability of the materials when the same adsorption and regeneration conditions are used repeatedly over an extended test period (days or weeks).

3.3.4 Laboratory-Scale Fluidized Bed Reactors

Jet-cup attrition testing was conducted through an outside laboratory. For the test, a sample of solid particles is contained in a small cup and high velocity gas is introduced to the cup tangentially. For the tests reported herein, the 20°C (70°F) gas was introduced at 300 ft/sec for two hours. Particle size distribution was measured using a Sympatec Halos Particle Size Analyzer before and after the test.

Two cold flow models were used during this program. Initial testing was conducted by PSRI Inc. In one of their standard testing setups and provided information on solid handling characteristics and could be used to measure heat transfer in air.

A second cold flow model was designed to mimic some of the basic fluidization and conveying processes that were designed into the pilot scale unit. A photo of this model is shown in Figure 9. The second model was constructed with two stages and multiple cyclones to allow investigations into fluidization regimes expected in the pilot facility. The benefit of this model was that it allowed for systematic testing of sorbent material at a wide range of fluidization gas velocities, and was used for operator training.

Figure 9. Photo of ADA Cold Flow Process Model.

The key system characteristics that are affected by sorbent properties that can be evaluated using the cold-flow models include:

- Fluid bed density
- Gas velocity and particle size required to achieve the formation of small bubbles
- Entrainment rate
- Heat transfer coefficient
- Sorbent size distribution
- Geldart classification
- 90% size distribution
- Particle density
- Bulk density (fluffed)
- Minimum fluidization velocity
- Heat of reaction

3.4 Selection of Sorbent Type

Prior to the commencement of DE-FE0004343, extensive sorbent screening had already been completed and many different types of sorbents were characterized. Two sorbent types were under consideration for this project: supported amines and activated carbons. The high-level pros and cons for each sorbent type were as follows.

- Supported Amines
  - Advantages
    - Higher CO_2 delta loading
    - Demonstrated performance in high moisture environments
  - Concerns
    - Effect of flue gas constituents (SO_2 was specifically a concern, similar to aqueous amine CO_2 capture systems)
    - Some, although not all, supported amines were not cyclically stable
    - Sorbent cost
- Activated Carbons
  - Advantages
    - Demonstrated performance in high moisture environments – a small decrease in CO_2 capacity was observed due to moisture, but for many materials the effect was not dramatic
    - Not chemically affected by the presence of SO_2 and other flue gas contaminants
    - Consistent cyclic stability
    - Acceptable sorbent cost
  - Concerns
    - CO_2 delta loading
    - Selectivity and CO_2 purity

For the purposes of final sorbent selection, a CO_2 delta loading was required. This CO_2 delta loading dictates the amount of sorbent that would be required to capture 90% of CO_2 from flue gas. The actual conditions (i.e., temperature and CO_2 partial pressure) to which the sorbent will be exposed are determined by the capture/regeneration process.

Heat for regeneration can be provided indirectly through heating coils or directly through steam or other heated gas. Any H_2O mixed with CO_2 leaving the regenerator will necessitate the addition of condensers to separate the H_2O from the CO_2 exhaust.

It should be noted that if steam is used for direct heating during regeneration, lower CO_2 partial pressure in the regenerator will result in a larger CO_2 working capacity. However, indirect heating allows for better control of the system to prevent hot-spots and potential damage to the sorbent as a result of exposure to high temperatures. The thermal capacitance of the steel steam tubes acts as a buffer between high temperature steam excursions and the isothermal regeneration of the sorbent. If the sorbent is exposed to repeated or excessive incidental temperature excursions higher than 120 °C, the amines may be off-gassed and the sorbent base structure irreparably damaged.

Additionally, direct contact heating of the sorbent with steam could result in a greater percentage of water adsorption and absorption by the sorbent, especially during transient operating conditions. Any water vapor that condenses within the system would be absorbed by the sorbent in a liquid form and physisorbed in vapor form. The subsequent heating of the sorbent in the regenerator would include heating of the water, which has a higher specific heat than the dry sorbent. The regenerator would require greater amounts of heat transfer to reach the same temperatures due to the entrained water. The additional moisture in the sorbent would create an energy penalty contradictory to a key goal of dry sorbent technology – to reduce the energy penalty of liquid sorbents caused by the latent and sensible heat required by the water component of the sorbent.

Another attribute associated with larger amounts of water absorption by the sorbent is the flowability. As the sorbent absorbs water, the adhesion of the sorbent increases, causing greater flow and transport difficulties within and between the fluidized bed systems. Dry sorbent requires lower gas pressure to transport and fluidize due to lower density, and has a lower likelihood of clogging transport lines. Higher density, moist sorbent would result in a higher energy penalty and a greater safety risk during operation due to higher required duct pressures.

The physical equipment and real-world operations of CO_2 capture would include startup and transient operating regimes that could result in large quantities of water condensation in the sorbent. The theoretical benefit of direct contact heating is not great enough to warrant risking unsustainable operating characteristics or the safety of personnel.

For two of the most promising activated carbon samples, the CO_2 capacity was measured using the TGA at temperatures between 30 to 120°C at three different partial pressures of CO_2, 0.081, 0.5, and 0.81 bar. Note that the TGA was operated at atmospheric pressure at the testing location where the ambient pressure was only 0.81 bar. Therefore, the 0.81 bar CO_2 partial pressure corresponds to 100% CO_2 gas.

For the sorbent selected for the 1 MWe pilot it is important to note that the H_2O concentration in the regenerator exhaust was a approximated to be 10 to 20 vol%. Thus, a CO_2 pressure of 0.8 to 0.9 bar was appropriate for regeneration. Therefore, for the purpose of comparing different sorbent types, the general conditions of 40°C, and 0.081 bar CO_2 for adsorption and 120°C and 0.81 bar CO_2 for regeneration were appropriate to calculate CO_2 delta loading.

Two different potential activated carbon sorbents were evaluated. The CO_2 loading versus CO_2 partial pressure for the two best activated carbons noted as Sorbent AM and Sorbent AN are provided in Figure 10 and Figure 11, respectively.

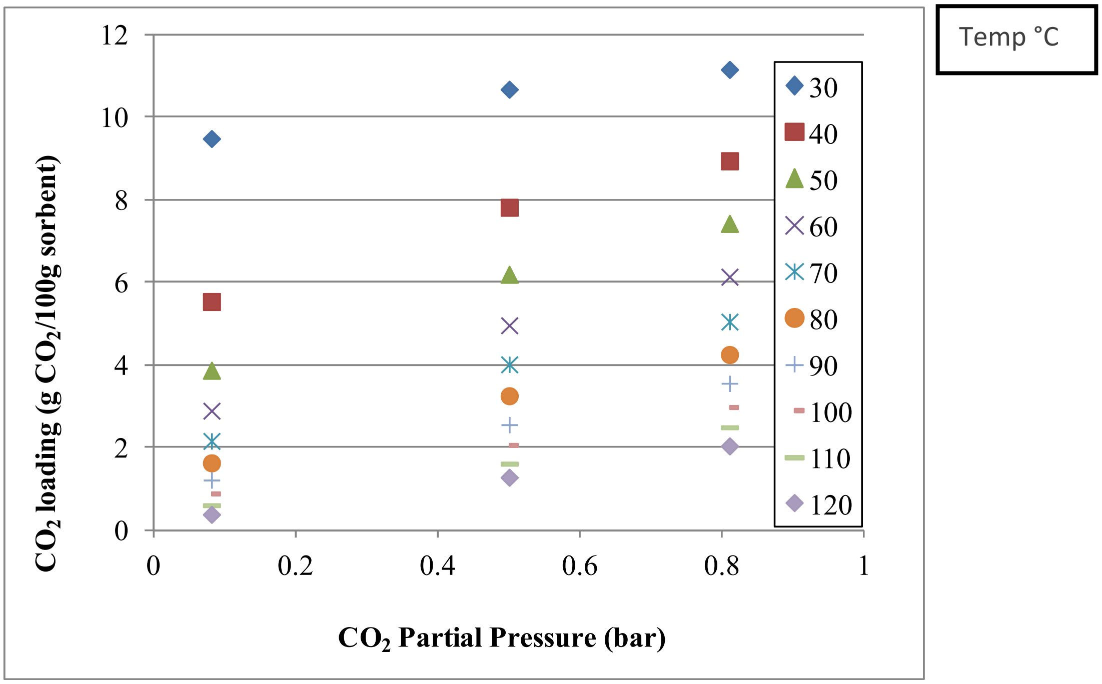


Figure 10. Sorbent AM (Activated Carbon) CO$_2$ Capacity at Various Temperatures and CO$_2$ Partial Pressures.

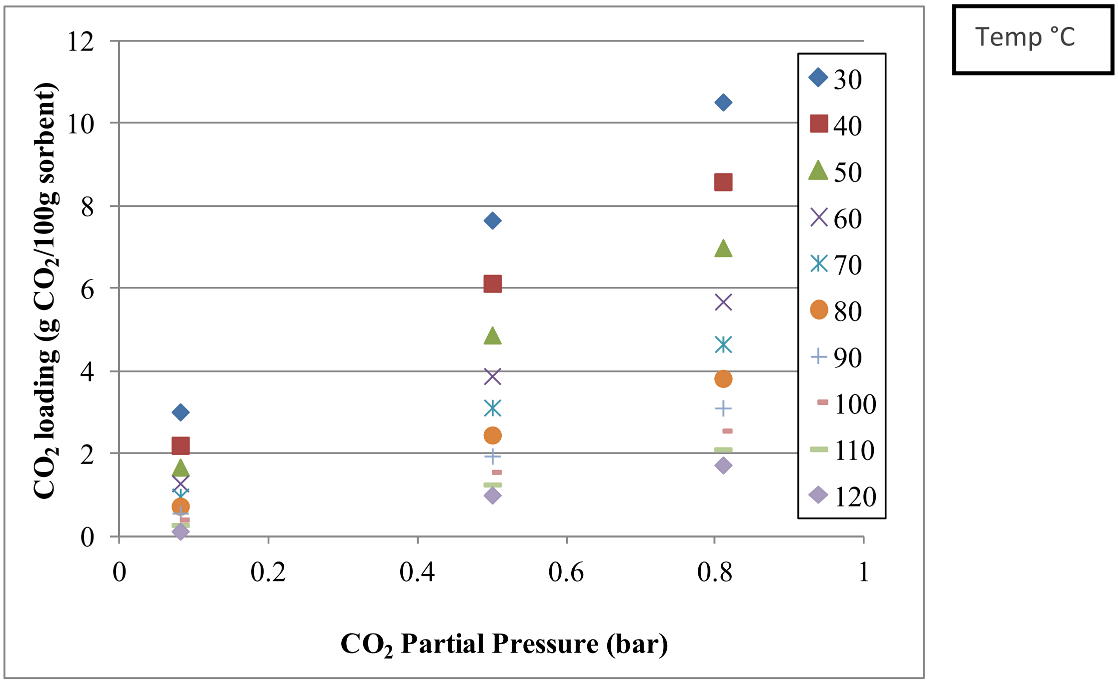


Figure 11. Sorbent AN (Activated Carbon) CO$_2$ Capacity at Various Temperatures and CO$_2$ Partial Pressures.

The relationship between the CO_2 loading and the CO_2 partial pressure for the activated carbon samples is indicative of what is often observed for physical adsorption samples. The CO_2 loading continues to increase with increasing CO_2 partial pressure, even at low temperatures. An important characteristic of the data shown in Figure 10 and Figure 11 is that the CO_2 loading can decrease dramatically when the sample temperature is increased.

Based on the loading values shown in Figure 10, the CO_2 delta loading of the first activated carbon sorbent, between the conditions of 40°C, and 0.081 bar CO_2 for adsorption and 120°C and 0.81 bar, was approximately 3.5 g CO_2/100 g fresh sorbent. Based on the loading values shown in Figure 11, the CO_2 delta loading of the second activated carbon sorbent, between the conditions of 40°C, and 0.081 bar CO_2 for adsorption and 120°C and 0.81 bar, was approximately 0.5 g CO_2/100 g fresh sorbent. Recall, the lower the CO_2 delta loading requires a higher sorbent circulation rate in the process in order to maintain CO_2 removal efficiencies. Higher circulation rates may result in larger equipment and higher capital costs which negatively impact the overall process economics.

The CO_2 loading at different temperatures and CO_2 partial pressures was also measured for three different supported amine sorbents. Sorbent F, the data for which is provided in Figure 12, is characterized by a tertiary amine grafted onto mesoporous silica. The total CO_2 uptake for the sorbent was higher than many other materials. However, for the process of CO_2 capture it is the CO_2 delta loading between adsorption conditions and desorption conditions that is more important. For the data shown in Figure 12, the delta loading, between the conditions of 40°C, and 0.081 bar CO_2 for adsorption and 120°C and 0.81 bar, is approximately 4.5 g CO_2/100 g fresh sorbent. The shape of the isotherms shown in Figure 12 indicated that the CO_2 uptake for this particular sorbent is due to both physical and chemical adsorption because there is a slight increase in the CO_2 loading at low temperature as the partial pressure is increased.

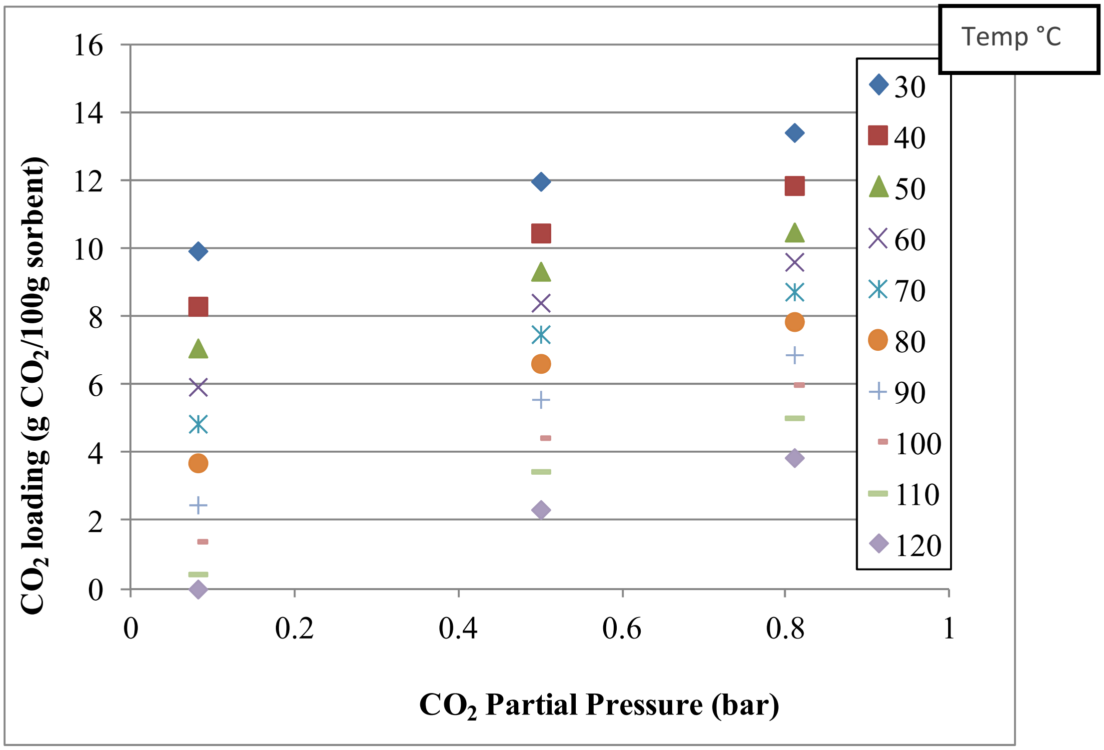


Figure 12. Sorbent F (Supported Amine) CO_2 Capacity at Various Temperatures and CO_2 Partial Pressures.

Sorbent CE, the supported amine sorbent used to generate the data shown in Figure 13, is characterized by a combination of polyethlyeneimine (PEI) and 3-(aminopropyl)triethoxysilane (APTES) immobilized on a porous silica support. For this material the CO_2 loading doesn't increase significantly between the 0.5 and 0.8 CO_2 partial pressure measurements at low temperature, which indicates that the CO_2 uptake is due primarily to chemical reaction, rather than physical adsorption. The delta CO_2 loading is calculated as approximately 4.0 g CO_2/100 g fresh sorbent.

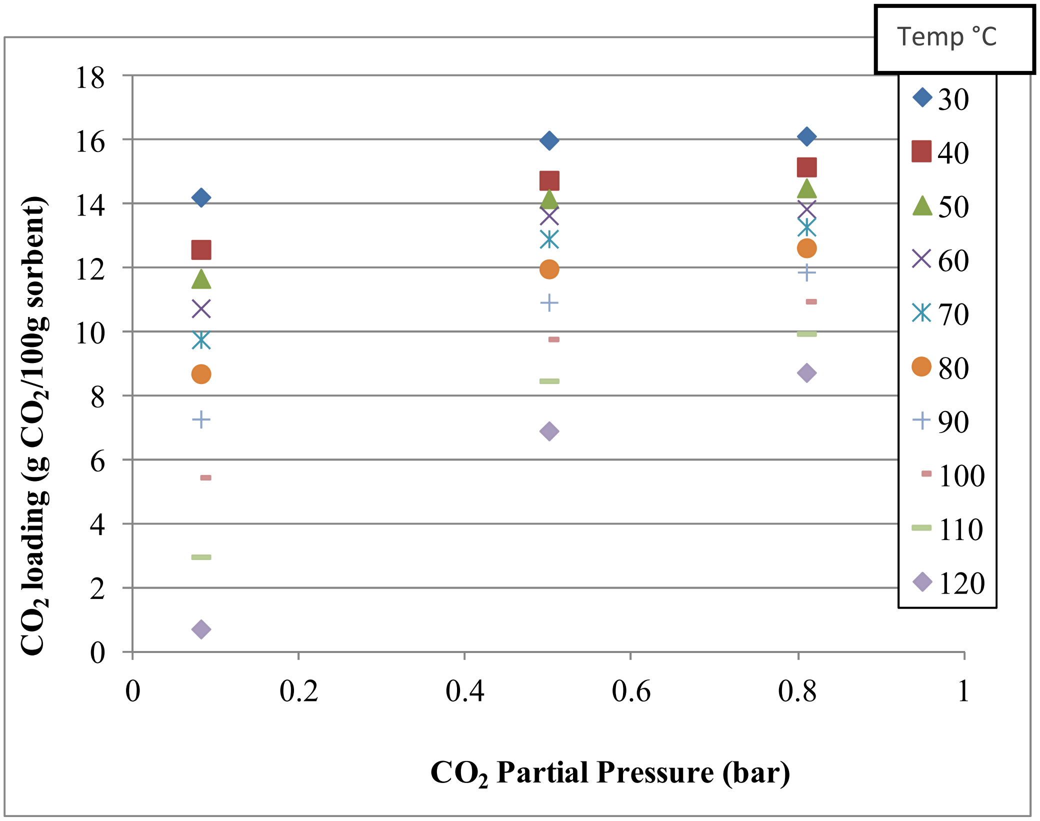


Figure 13. Sorbent CE (Supported Amine) CO_2 Capacity at Various Temperatures and CO_2 Partial Pressures.

Sorbent BN, the amine functionalized sorbent that was used to generate the CO_2 loading profiles provided in Figure 14, is characterized by a primary amine grafted onto the surface of an ion exchange resin. Similar to the loading data provided in Figure 13, the results shown in Figure 14 indicate a sorbent whose main CO_2 uptake can be attributed to a chemical reaction other than physical adsorption. For the data shown in Figure 14 the estimated delta CO_2 loading is approximately 6.0 g CO_2/100 g fresh sorbent under the aforementioned adsorption/regeneration conditions.

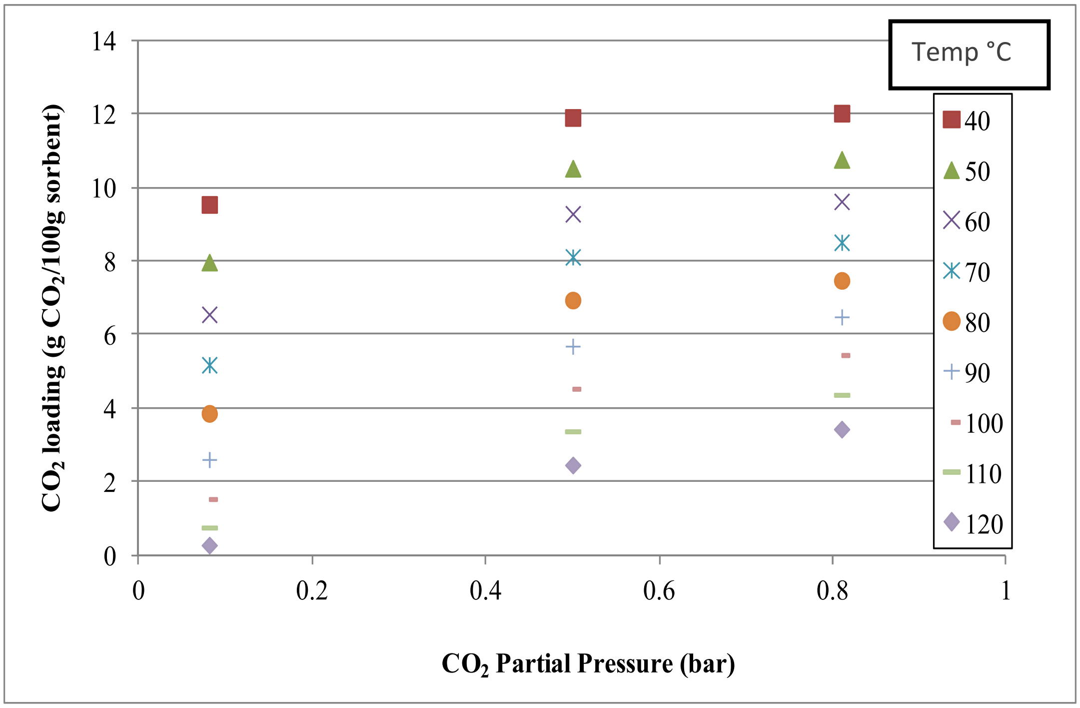


Figure 14. Sorbent BN (Supported Amine) CO_2 Capacity at Various Temperatures and CO_2 Partial Pressures.

One of the main reasons supported amine sorbents were under consideration is their greater working CO_2 capacity, which is a result of the highly exothermic reaction between CO_2 and the amine. When carrying out a TSA process, it is important to control the temperature during both adsorption and regeneration. Assuming an enthalpy of reaction of adsorption of -75 kJ/mol CO_2 the heat removal requirement can be easily calculated:

$$Q_{ads} = 0 = mCp\Delta T + \Delta H \qquad (12)$$

Solving for ΔT provides the increase in temperature due to the heat of reaction with CO_2. For an immobilized amine adsorbent with a working CO_2 capacity of 7 g CO_2/100 g fresh sorbent, an enthalpy of reaction of -75 kJ/mol, and a specific heat capacity of 1.0 J/g·K, the increase in temperature would be nearly 80°C. Assuming that the starting temperature was approximately 40°C, the adsorber would increase in temperature to over 120°C. However, the regeneration temperature for supported amine adsorbents is often in the range of 100 to 120°C (i.e., temperature swing of 60 to 80°C between adsorption and regeneration), so clearly this undesirable temperature increase resulting from the exothermic reaction with CO_2 would be prohibitive for a TSA process. Also, note that this temperature increase calculation is actually conservative because the heat generated by adsorbed H_2O was not included in this simplified calculation. Approximately the same magnitude of heat input would be required in the regenerator, where an endothermic reaction occurs. Therefore, it is equally important to have effective heat transfer during adsorption and regeneration. Note that the sensible heat has not been included and will also increase the cooling (adsorption) and heating (regeneration) requirements.

At the simplest level, the minimum specific energy can serve as a criterion, as discussed by Berger and Bhown[35] where the minimum specific theoretical energy for chemical and physical adsorbents was systematically modeled using Langmuir isotherms. They found that there was a minimum specific theoretical energy at a given regeneration temperature. This work reported that the minimum specific energy for adsorption-based CO_2 capture was attainable by an adsorbent with an enthalpy of adsorption of -64 kJ/mol CO_2 and a regeneration temperature of 160°C. Note that the minimum specific energy increased rapidly when the enthalpy of adsorption approached -25 kJ/mol CO_2 (approximately that for some physical adsorbents[36]). Although the minimum specific energy identified could theoretically be attained by a supported amine material (based on the enthalpy of adsorption) a regeneration temperature of 160°C is too high because it will lead to amine decomposition.[37] Fortunately, the minimum specific energy was relatively flat within the enthalpy range of -50 to -75 kJ/mol CO_2.

A summary of the delta loading of two activated carbon sorbents and three potential supported amine sorbents are shown in Table 6. Although one of the activated carbon sorbents could potentially be competitive with the supported amine sorbents regarding the high delta loading and associated regeneration heat duty, the material handling requirements would be significantly more challenging because the mass of material that was circulated in the system wouold need to be higher to achieve the same overall CO_2 removal. The costs associated with the increase in sorbent circulation rates negatively impacted the process economics. Therefore, supported amine sorbents were selected as the sorbent family which would be further investigated under this project.

Table 6. Comparison of Activated Carbon and Amine Sorbents.

	Activated Carbon		Supported Amine		
	AM	AN	F	CE	BN
Delta CO_2 Loading (g CO_2/100 g fresh sorbent)**	3.5	0.5	4.5	4.0	6.0

**Calculated using the data shown in Figures 9 through 11 with an adsorption condition of 40°C and 0.081 CO_2 partial pressure and regeneration conditions of 120°C and 0.81 CO_2 partial pressure.

3.5 Specific Sorbent Selection

When selecting a single supported amine sorbent to be characterized and used for the design basis of the 1 MWe pilot, several key properties were taken into consideration. The different sorbents and their properties are provided in Table 6. Note that some considerations, such as the ability to produce the material in appropriate quantities for the 1 MWe project, only apply to the current technical readiness, not to the future of the material to be produced in such quantities. Similarly, while it may be possible to utilize a sorbent with a mean particle diameter less than 50 µm in a commercial process, this was not the case for the processes being considered under this project, therefore a minimum particle diameter was established.

Table 6. Key Sorbent Selection Properties.

	Sorbent F	Sorbent CE	Sorbent BN
Can Be Produced at Required Scales	No	Yes	Yes
Particle Size Greater than 50 μm	No	Yes	Yes
Stable Under High Moisture Conditions*	Yes	Questionable	Yes
Stable at 120°C	Yes	Questionable	Yes
Delta CO_2 Loading ($g\ CO_2$/100 g fresh sorbent)**	4.5	4.0	6.0

*Based on the amine application technique and public literature

**Calculated using the data shown in Figure 12, Figure 13, and Figure 14 with an adsorption condition of 40°C and 0.081 CO_2 partial pressure and regeneration conditions of 120°C and 0.81 CO_2 partial pressure.

Comparing the three different amine functionalized sorbents listed in Table 6, the aminated ion exchange resin (sorbent BN) was selected as the sorbent that would be further characterized and used as the basis for the design of the 1 MWe pilot.

3.6 Sorbent Characterization and Analysis

3.6.1 CO_2 Delta Loading

Since one of the key selection criteria for the sorbents was the CO_2 delta loading, the CO_2 loading at different partial pressures was measured. However, after the aminated ion exchange resin was selected for the 1 MWe pilot design basis full isotherms were required. In addition to the CO_2 partial pressures of 0.081, 0.5, and 0.81 (gas blends of 10%, 61%, and 100% CO_2 by volume, respectively) previously measured, additional gas blends were used to measure the CO_2 uptake at several temperatures and 0.4 and 0.15 CO_2 partial pressure (gas blends with 4.9% CO_2 and 18.5% by volume, respectively). Full isotherms were developed using a Langmuir fit, which was acceptable based on the shape of the isotherms for this particular sorbent. Due to cooling water restrictions, it was determined that the sorbent would not be fully characterized at temperatures below 40°C. The measured and calculated loading at different partial pressures and temperatures are provided in Figure 15.

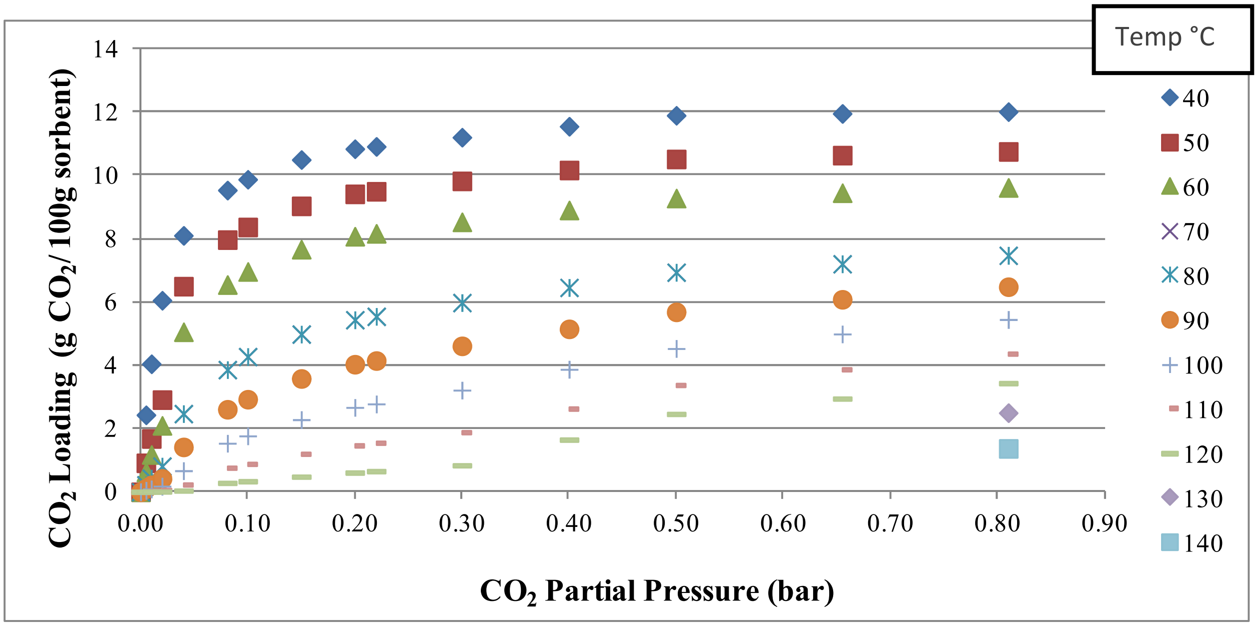


Figure 15. CO_2 Loading at Different Temperatures for Selected Sorbent (BN).

The CO_2 delta loading must be determined by considering the process in which the sorbent will be used. Based on the CO_2 capture process that was developed, the CO_2 capture occurred at approximately a CO_2 partial pressure of 0.15 bar and temperature of 40°C, which corresponds to a CO_2 loading of approximately 10.5 g CO_2/100 g fresh sorbent. The regeneration conditions were varied during the 1 MWe test, but as a general estimate during sorbent development the regeneration CO_2 partial pressure was assumed to be 0.8 bar and the temperature was assumed to be 120°C, which corresponds to a CO_2 loading of 3.5 g CO_2/100 g fresh sorbent. Therefore, for this sorbent under the assumed operating conditions, the delta CO_2 loading is 7.0 g CO_2/100 g sorbent. Increasing the delta loading will likely result in cost reductions for the overall CO_2 capture process. However, it is important to note that this delta CO_2 loading is superior to that currently exhibited in larger aqueous monoethanol amine CO_2 capture systems (estimated to be approximately 4.5 g CO_2/100 g fresh solvent)[1].

3.6.2 Cyclic Stability

Although the commercial-scale cost of most potential CO_2 sorbents has yet to be determined, it is realistic to believe that sorbent costs will be greater than that of aqueous solvent costs per unit mass. Therefore, it is important that any CO_2 sorbent can be reused for thousands of adsorption/regeneration cycles. The major concern regarding the cyclic stability is related to degradation of the amines due to high temperatures. To evaluate the temperature stability of the aminated ion exchange resins, five adsorption/regeneration cycles were completed at six different regeneration temperatures. The adsorption temperature was 50°C for all cycles (this test was completed before the adsorption temperature of 40°C had been selected for the process). Five adsorption/regeneration cycles were completed at each test condition, which was only enough to indicate significant issues with the temperature stability. For all the tests, the CO_2 loading was measured with 100% CO_2 (i.e., PCO_2 = 0.81 bar). Therefore the delta loading provided in Figure 16 should not be considered the CO_2 delta loading under the actual process conditions.

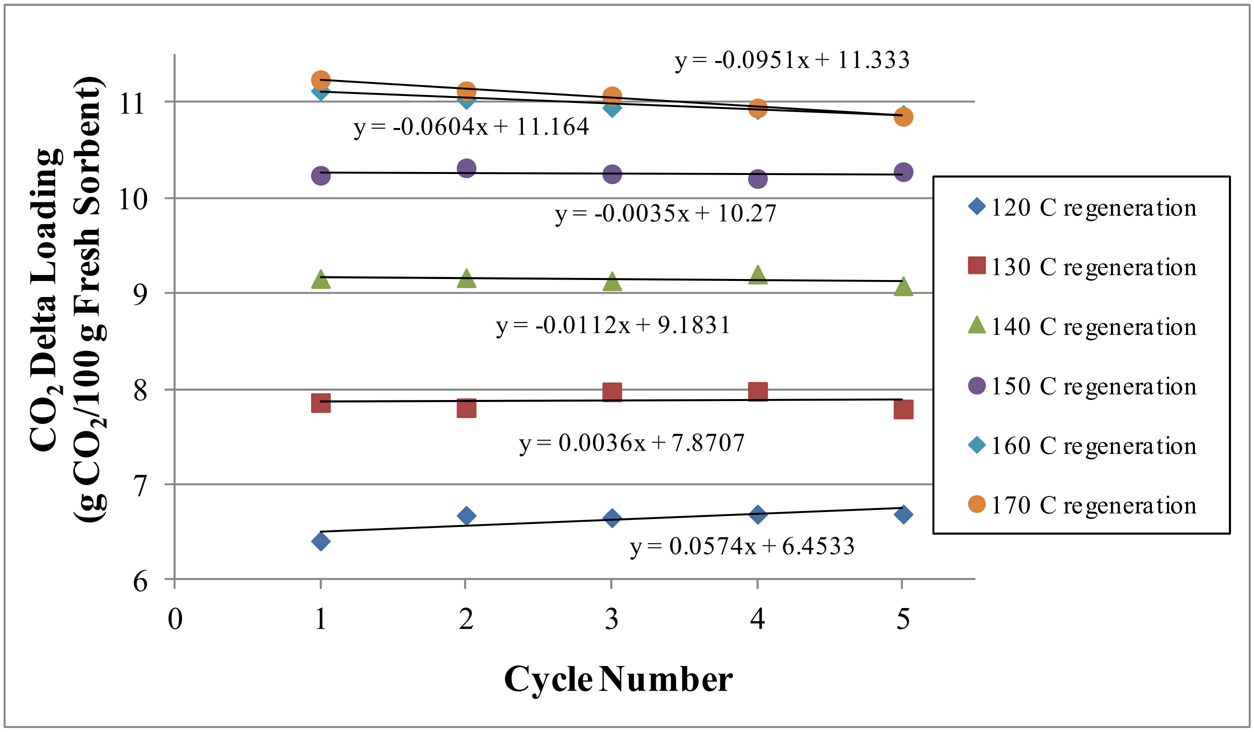


Figure 16. CO_2 Delta Loading Under Pure CO_2 Atmosphere for Selected Sorbent (BN).

The results from each test were fit with a linear curve and the equation for each curve is also provided on Figure 16. Every temperature above 130°C exhibited a negative slope, indicating a loss in CO_2 delta loading. The negative slope calculated when regenerating at temperatures less than 160°C was more severe than compared to lower regeneration temperatures. Based on the data shown in Figure 16, it was decided that a longer test would be completed cycling between temperatures of 50°C and 130°C under a pure CO_2 atmosphere. A pure CO_2 atmosphere was used to decrease the time required for gas switching if the actual adsorption partial pressure of CO_2 was used. The CO_2 delta loading recorded during this test is provided in Figure 17.

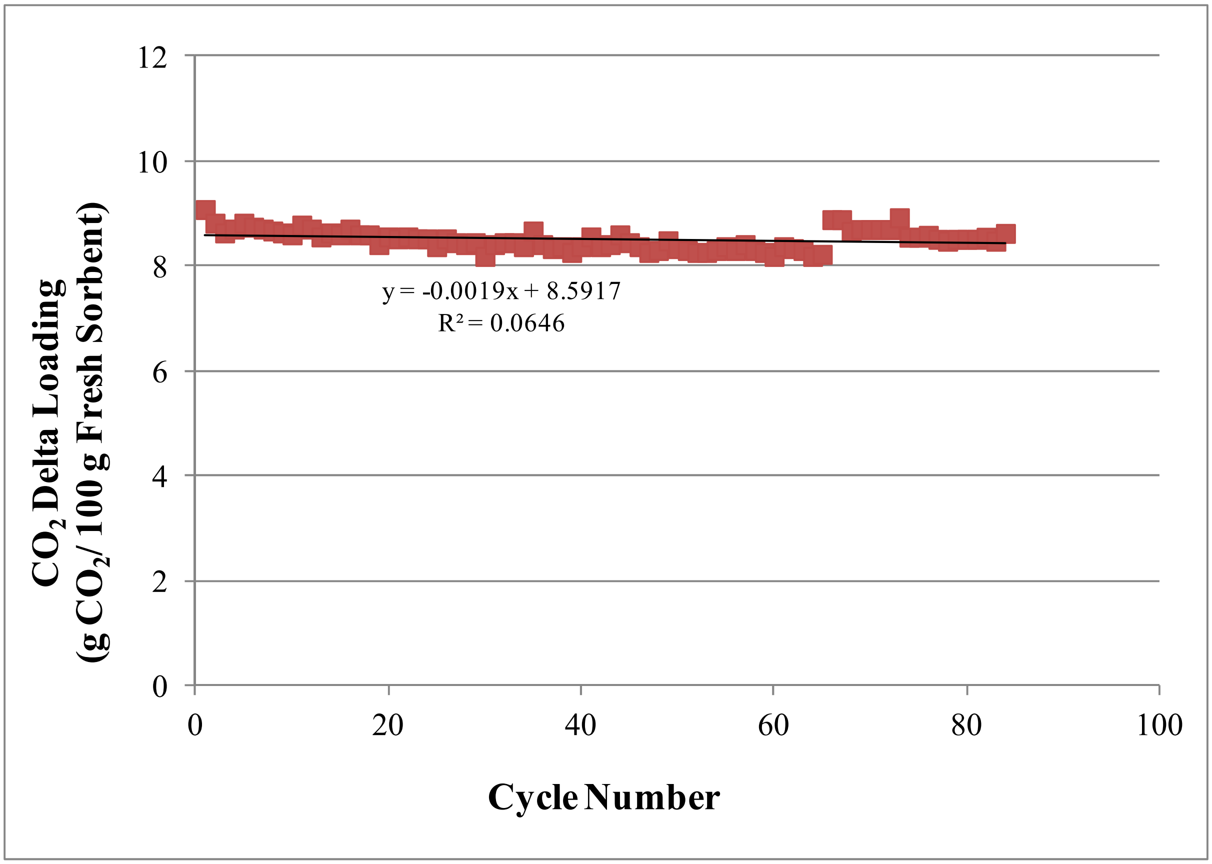


Figure 17. CO_2 Delta Loading Measured Cycling Between 50°C and 130°C with 100% CO_2 for Selected Sorbent (BN).

Note that the CO_2 delta loading is larger than the estimated delta loading previously discussed because the data shown in Figure 17 was collected under a pure CO_2 atmosphere (i.e., the partial pressure during the low temperature steps was 0.81 bar, but the CO_2 partial pressure during adsorption is closer to 0.15 bar). The CO_2 delta loading during this test decreased slowly as the cycle number increased until a jump in the CO_2 delta loading was observed at approximately the sixty-fifth cycle. At this cycle the TGA program had reached its maximum number of steps and was restarted. The nitrogen flush that is used at the beginning of every TGA test was inadvertently repeated when the program was restarted. When the nitrogen flush occurred the sorbent regained its original CO_2 delta loading. Therefore, it was determined that the pure CO_2 atmosphere (which was used to decrease the test duration, not because it is indicative of realistic process conditions) could be leading to a bias of the results.

Due to the altitude at ADA's laboratory in Colorado (~5,800 feet) and the use of an atmospheric pressure TGA instrument, the maximum CO_2 partial pressure used for investigating loading after regeneration was only 0.81 bar. However, in practice, the pilot was located at an elevation of approximately 430 feet above sea level and the regenerator was pressurized by a blower creating the necessary motive force to fluidize the bed of sorbent particles. This resulted in a partial pressure of nearly 1 bar CO_2 as well as 0.1 bar moisture, oxygen, and nitrogen in the regenerator in the pilot plant. Additional tests were conducted with sorbent BN to assess performance through a partial pressure of 1 bar. These data are shown in Figure 18. The results at 1 bar are very similar to the results at 0.8 bar.

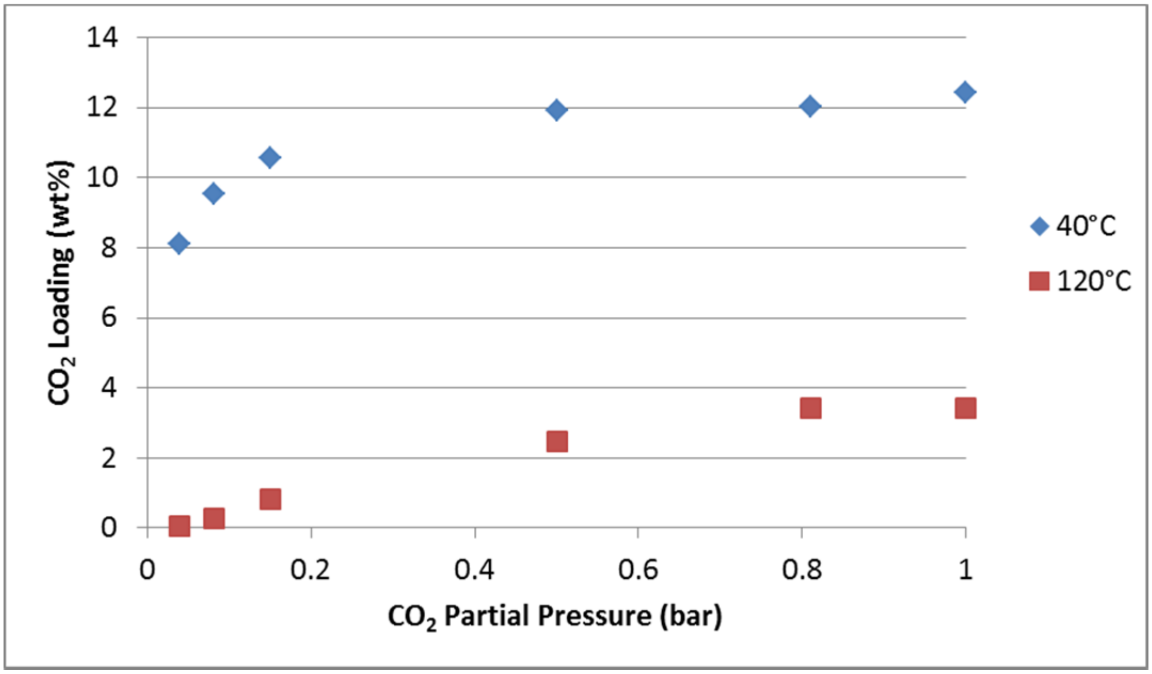


Figure 18. Sorbent BN Isotherms through 1 bar.

3.6.3 Reaction Kinetics

Assessments of kinetics were evaluated in the laboratory using data from the counter-flow (trickle down) reactor tests and from TGA tests.

The trickle-down reactor data was used to determine the approximate height of the adsorber and the rate at which CO_2 was adsorbed within the height limitations of the process laboratory. A series of tests were conducted to simulate CO_2 loading on the sorbent as it passes through a counter-current adsorber of unknown height. This was done by simulating a flue gas stream at the bottom of the apparatus adsorber that would result in a CO_2 concentration of 1.2% at the top of the adsorber section while feeding sorbent (i.e., what would be observed at the top of a commercial system where 90% of the CO_2 had been removed). Following this test run, the sorbent collected at the bottom of the adsorber was moved back to the feeder hopper and the flue gas CO_2 concentration was increased so that the outlet concentration equaled the inlet concentration of the previous test run. This was repeated several times until the inlet concentration at the bottom of the adsorber reached a concentration of 10.3%. This approximated the adsorber height required to reduce the CO_2 concentration from 10.3% entering the bottom of the adsorber to 1.2% exiting the top of the adsorber, if fresh sorbent was introduced at the top of the adsorber. In reality, the sorbent introduced at the top of the adsorber would already have some CO_2 adsorbed since it will not be fully regenerated.

A series of parametric tests were conducted initially to establish conditions for the kinetic evaluation test that represented good sorbent distribution throughout the reactor with minimal particle tracking. A summary of the particle velocity compared to the gas velocity from testing in shown in Figure 19. For the kinetic tests, the flue gas velocity was maintained at 1.68 m/s (5.5 ft/sec) and the sorbent feed rate was maintained at 0.45 kg/min (1 lbs/min). This combination was chosen since the sorbent was well distributed at this velocity and the feed rate was based on a nominal 5% working capacity of the sorbent, based on isotherms measured using the TGA at the process conditions of the trickle-down reactor.

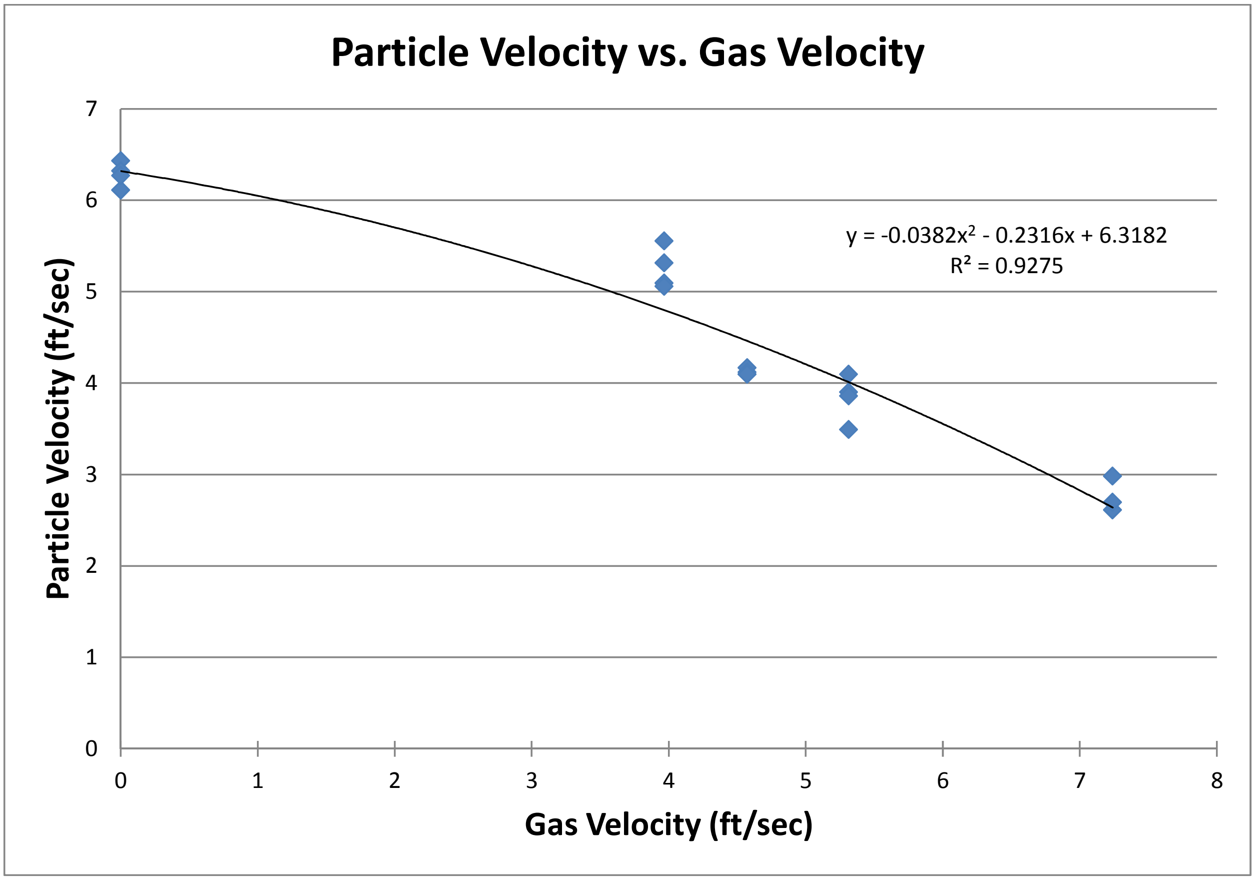


Figure 19. Summary of Gas Velocity Compared to Particle Velocity in the Trickle-Down Reactor.

A summary of the data collected during kinetic testing is presented in Table 7, where run 1 conditions were indicative of the top of the adsorber where the regenerated sorbent enters and the CO_2-lean gas exits and run 10 conditions represented the bottom of the adsorber where loaded sorbent is removed, and high-concentration flue gas enters. Between runs 7 and 8, the sorbent was heated in an oven under a CO_2 blanket overnight to bring the sorbent temperature back up. Unfortunately, it appears the sorbent was regenerated due to the high CO_2 removal experienced during run 8 and the high temperatures in the adsorber section. Thus, the total CO_2 loading after runs 8, 9, and 10 are not necessarily indicative of the total CO_2 loading on the sorbent expected at the outlet of a full-scale trickle-down reactor.

Although the data from run 8 was discounted as an indicator of cumulative CO_2 loading on the sorbent after repeated exposure to increasing concentrations of CO_2 (i.e., runs 1 through 7), the test run provided key insights into the reaction kinetics of a regenerated sorbent when exposed to CO_2 in a gas stream at a temperature and concentration representative of post-combustion capture. Based upon test conditions, the average velocity of particles in the system is expected to be approximately 1.2 m/s (4 ft/s). Recall that the adsorber section was 2.29 m (7.5 ft) long. Thus, the particles were exposed to CO_2-laden gas for nominally 1.9 seconds for each run. During run 8, the sorbent adsorbed 2.6g CO_2/100g sorbent. The implications of this result are discussed in more detail in Section 6.3.2.

Table 7. Adsorber Height and Reaction Kinetics Test Data Summary from the Trickle Down Reactor.

Test Run	Inlet CO_2 Conc. (%)	Outlet CO_2 Conc. (%)	Change in Sorbent Loading During Run (wt CO_2/sorbent, %)	Sorbent Loading at Outlet of Bed (wt CO_2/sorbent, %)	Sorbent Temp. (°C) [(°F)]
1	3.0	1.2	1.6%	1.6%	40 [105]
2	4.5	3.0	1.3%	2.9%	47 [117]
3	5.5	4.5	0.9%	3.8%	43 [109]
4	6.7	5.6	1.0%	4.8%	45 [113]
5	7.2	6.6	0.5%	5.3%	41 [106]
6	8.1	7.4	0.6%	5.9%	37 [99]
7	8.6	8.1	0.4%	6.4%	37 [99]
Heating and CO_2 Treatment of Sorbent Overnight					
8	12.5	9.5	2.6%	NA	37 [99]
9	9.2	8.6	0.5%	NA	42 [107]
10	10.3	9.0	1.1%	NA	42 [107]

One of the key limitations of this test is maintaining the CO_2 loading on the sorbent between runs. For example, the sorbent may continue to adsorb CO_2 from the gas in the interstitial spaces while it is collected in bulk at the bottom of the reaction column, even if the temperature is maintained. Noting these limitations, extrapolating results from the applicable runs suggest that a total of 27 m (90 ft) of adsorption height should result in 90% CO_2 capture at a flue gas velocity of 1.68 m/s (5.5 ft/sec). Interstage cooling will also be required and is not taken into account in this projected height. Considering the velocity of particles in the system is expected to be approximately 1.2 m/s (4 ft/s), a 27m high adsorber represents less than 25 seconds of sorbent residence time to achieve 90% CO_2 capture at conditions representative of post-combustion capture (40°C and inlet CO_2 of 10.3%).

Results from TGA testing can provide more quantitative information on the kinetics of adsorption and desorption. Figure 20 displays the mass of beaded sorbent BN during experiments where the temperature (shown as a green line in the figures) was cycled. The weight of the sample (red line in the figures) is also shown. Note that the CO_2 concentration was not varied during these tests; the entire test was completed in a pure CO_2 atmosphere. Only the temperature was changed between 40°C and 120°C during each cycle. The temperature change from 40°C to 120°C or from 120°C to 40°C required about 10 minutes, due to limitations of the apparatus.

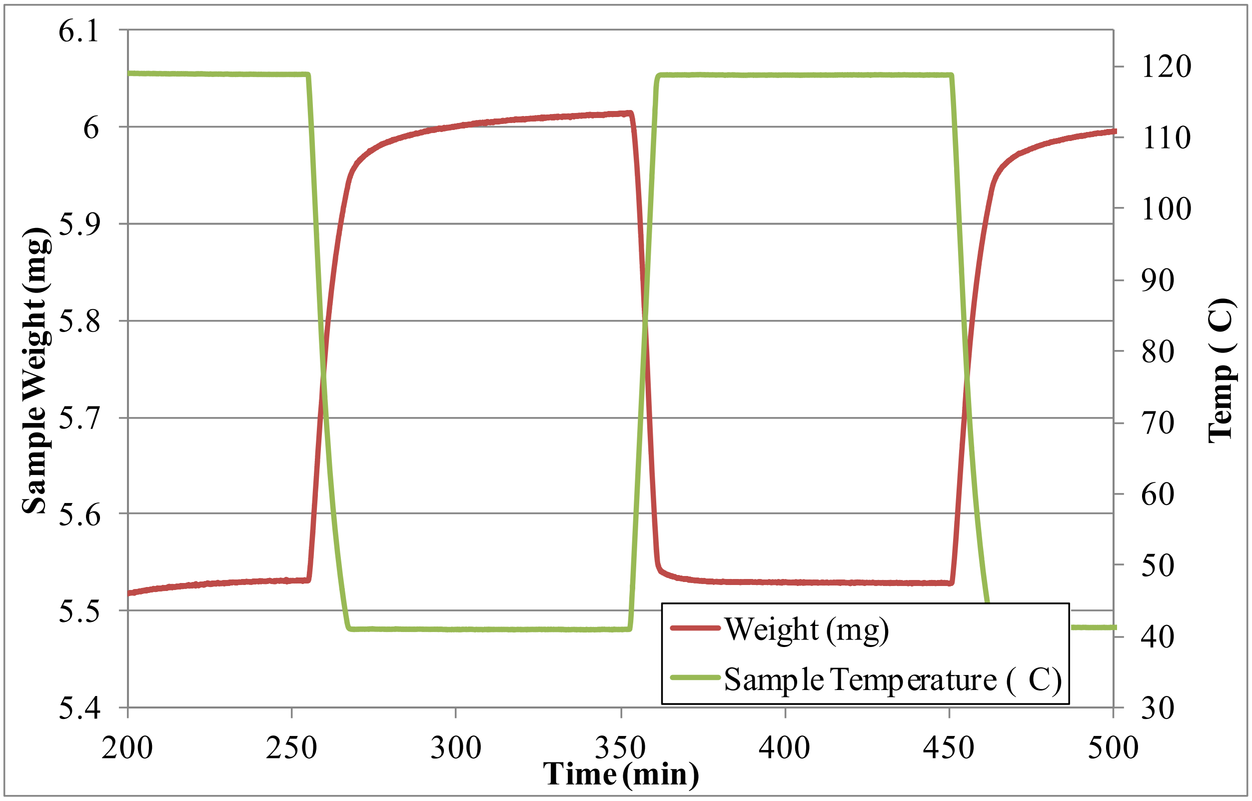


Figure 20. Sample Weight and Temperature during Rapid Heating and Cooling for Sorbent BN in 100% CO_2 atmosphere.

These results do not represent conditions expected in the pilot-scale fluidized beds, because of the relatively long time required to change temperature (circa 10 minutes) and the constant CO2 partial pressure of 0.81 bar. Nevertheless, during adsorption in the pure CO_2 atmosphere of the TGA, the kinetics are fast initially with nominally 90% or more of the CO_2 loading occurring at nominally the same rate as the change in temperature. At a regeneration temperature of 120°C, the CO_2 release is fast. The weight of the sorbent stabilizes in less than 10 minutes as the TGA temperature reached the setpoint. This data indicates that the CO_2 release in the regenerator will be fairly rapid.

Additional TGA testing was conducted with Sorbent BN at cycle times between adsorption and desorption more representative of potential pilot conditions in both time at each temperature, and partial pressure of CO_2 at each condition, to provide a better indication of potential reaction kinetic limitations. The TGA was loaded with fresh sorbent and cycled between 120°C, 0.81 bar and 40°C, 0.15 bar for 74 cycles. Figure 21 shows the initial time period of the experiment. The initial desorption cycle was longer to "condition" the as-received sample. The temperature ramp from 120°C to 40°C required 12 minutes, and the temperature ramp from 40°C to 120°C required 8 minutes, which was a limitation of the apparatus.

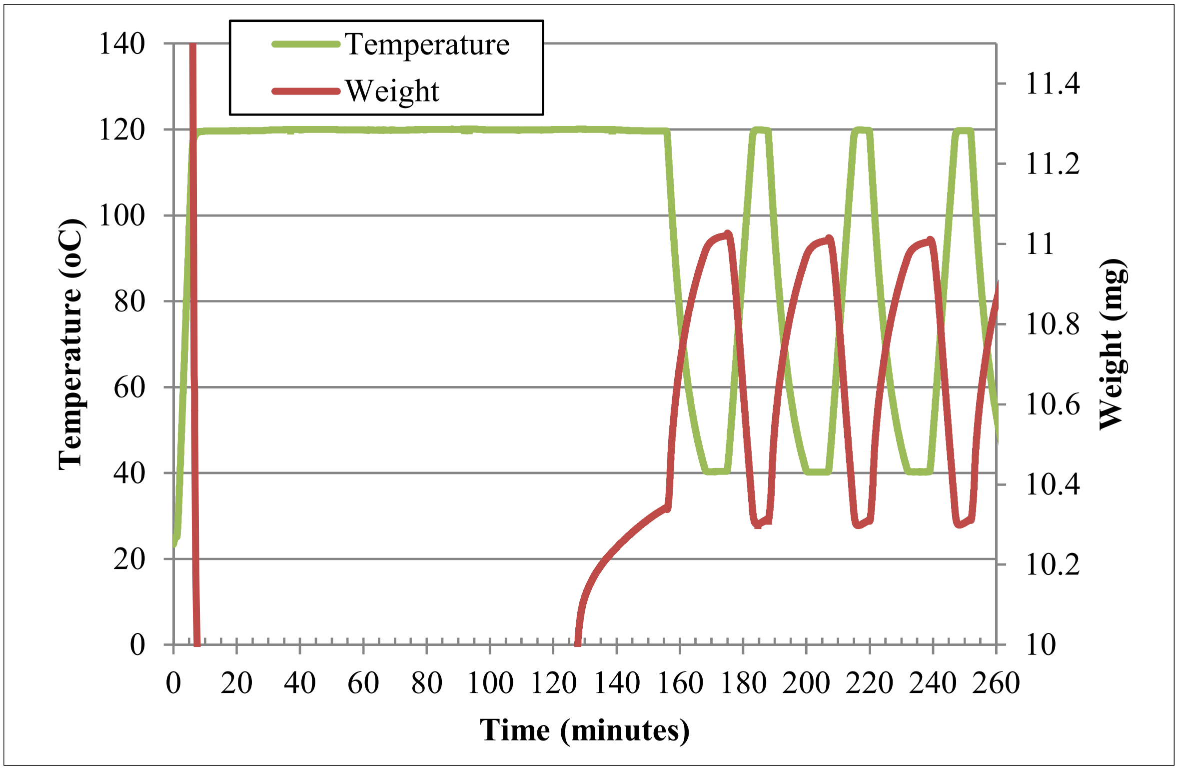


Figure 21. Sample Weight and Temperature When Cycling Sorbent BN between 40°C/0.15 bar CO_2 and 120°C/0.8 bar CO_2.

Selected results from the initial and final cycles (Table 8) indicate that after the initial desorption cycle, the time required to reach 90% of equilibrium capacity was typically less than 11 minutes on the adsorption cycle and nominally 7 minutes on the desorption cycle. The delta CO_2 loading and kinetics of adsorption and desorption were consistent among the cycles (after the initial cycle). Although conditions during this test were more representative of the pilot-scale fluidized beds, the relatively long heating and cooling times in the TGA make it difficult to apply these results quantitatively to the pilot-scale reactor. The TGA results represent an upper bound on the expected time to reach equilibrium capacity, because temperature can be changed more quickly in the fluidized beds.

Table 8. Multiple Cycle TGA Test of Sorbent BN.

Condition	Cycles	Time to 80% Capacity Equilibrium (minutes)	Time to 90% Capacity Equilibrium (minutes)	CO_2 Loading (g CO_2/ 100g Fresh Sorbent)	Delta Loading (g CO_2/ 100g Fresh Sorbent)
Desorption	1	17.46	22.78	5.22	
Adsorption		8.27	10.69	12.14	6.92
Desorption	2	6.26	6.98	4.96	
Adsorption		7.30	9.96	12.03	7.07
Desorption	3	5.77	6.74	4.89	
Adsorption		8.16	11.30	11.98	7.08
Desorption	72	7.08	7.83	5.08	
Adsorption		8.29	10.77	11.78	6.71
Desorption	73	6.91	7.08	5.06	
Adsorption		8.47	10.95	11.77	6.72
Desorption	74	7.10	7.79	5.06	
Adsorption		8.75	11.05	11.77	6.72

The results from the trickle-down reactor tests suggest that the adsorption kinetics are very fast, which is consistent with TGA results. This is a critical insight for sorbent-based capture systems. Specifically, if the system is designed to operate within the region of very fast kinetics, the residence time within the adsorber and the resulting size and capital costs of the adsorber can be optimized. Under typical cycling conditions, the rate of desorption, is faster than the rate of adsorption, which has implications for the design of the adsorber and desorber.

3.6.4 Effect of Flue Gas Constituents

Previous work indicated that SO_2 will form heat stable salts with the amine functional groups of any supported amine sorbent. As a result, potential process options to reclaim chemically fouled sorbent back to its original chemical makeup were investigated and led to patent US 2014-0079612 A1.

Prior to sorbent selection, all potential CO_2 sorbents were evaluated with moisture in simulated flue gas. Therefore, it was already known that sorbent BN could adsorb CO_2 in the presence of moisture. However, the moisture uptake had not been previously measured on this particular sorbent. The moisture uptake of a sorbent is important because it could affect the energy penalty and process design. If a sorbent adsorbs moisture during the adsorption step of the CO_2 capture process and then releases it during the regeneration step, the sorbent will have an effective H_2O working capacity. For the H_2O to be released during the regeneration step, energy must be added equivalent to the enthalpy of vaporization, which will in turn become part of the regeneration energy. It is important to assess the moisture uptake so this added energy penalty can be quantified and minimized.

The TGA/MS setup shown previously was utilized to assess the moisture loading on the aminated ion exchange resin at different temperatures and moisture levels in the feed gas. First the sorbent was dried by sending dry N_2 from a compressed gas cylinder directly to the TGA. This drying step was necessary to achieve an initial weight of the sorbent. The moisture was introduced into the TGA by diverting 40% of the TGA feed gas through a temperature-controlled bubbler. The bubbler temperature set point was initially room temperature, approximately 22°C (72°F). After the weight stabilized, the temperature of the bubbler was slowly increased, which also increased the concentration of H_2O in the gas stream. A mass spectrometer was used to measure the moisture concentration in the TGA exhaust gas. The entire weight uptake observed during these tests was assumed to be due to moisture because there was no CO_2 in the feed gas. The results of the moisture uptake experiment are provided in Figure 22.

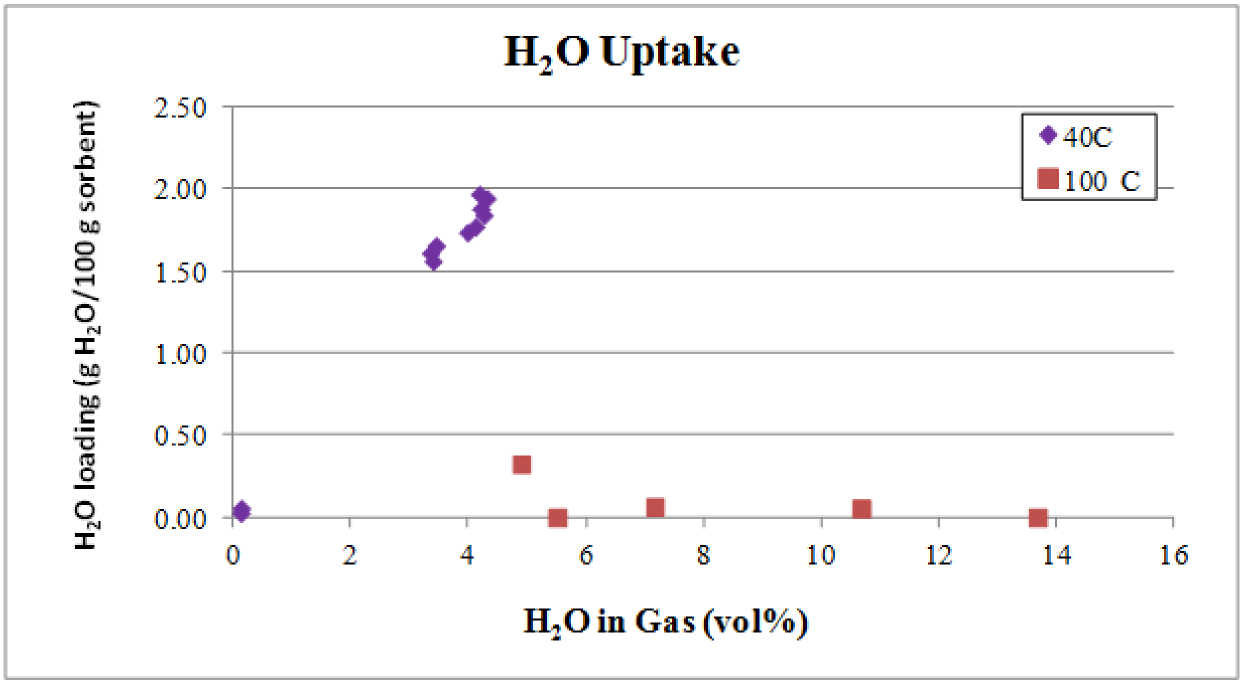


Figure 22. Moisture Uptake Data for Sorbent BN.

One important conclusion drawn from the data provided in Figure 22 was that the relationship between moisture loading and the moisture concentration in the gas was nearly linear at 40°C. In addition, the tests showed that the moisture loading at or above 100°C was negligible. From this information, the delta moisture loading can be estimated for the sorbent under the process conditions. Based on the current process conditions of adsorption at 40^0 C, and a process gas at 40° and 1.5 bar, it is expected that the delta moisture loading will be approximately 1.0 g H_2O/100 g fresh sorbent. This is an important conclusion because the adsorption/release of moisture will result in additional heat removal/input, respectively, which must be taken into account in the equipment design.

A second test was conducted to assess the moisture uptake on sorbent BN without the presence of CO_2. A gas stream of N_2 with 5.7% H_2O was used, although these conditions are slightly more humid than the process being considered. The results of the moisture uptake are provided in Figure 23. The moisture uptake increased dramatically at lower temperature. For example, at 40°C the moisture uptake was over 2 g H_20 / 100 g sorbent, while it was approximately 0.7 g H_20 / 100 g sorbent at 60°C. As expected, when the sorbent temperature was greater than or equal to 100°C, the moisture content in the sorbent was negligible.

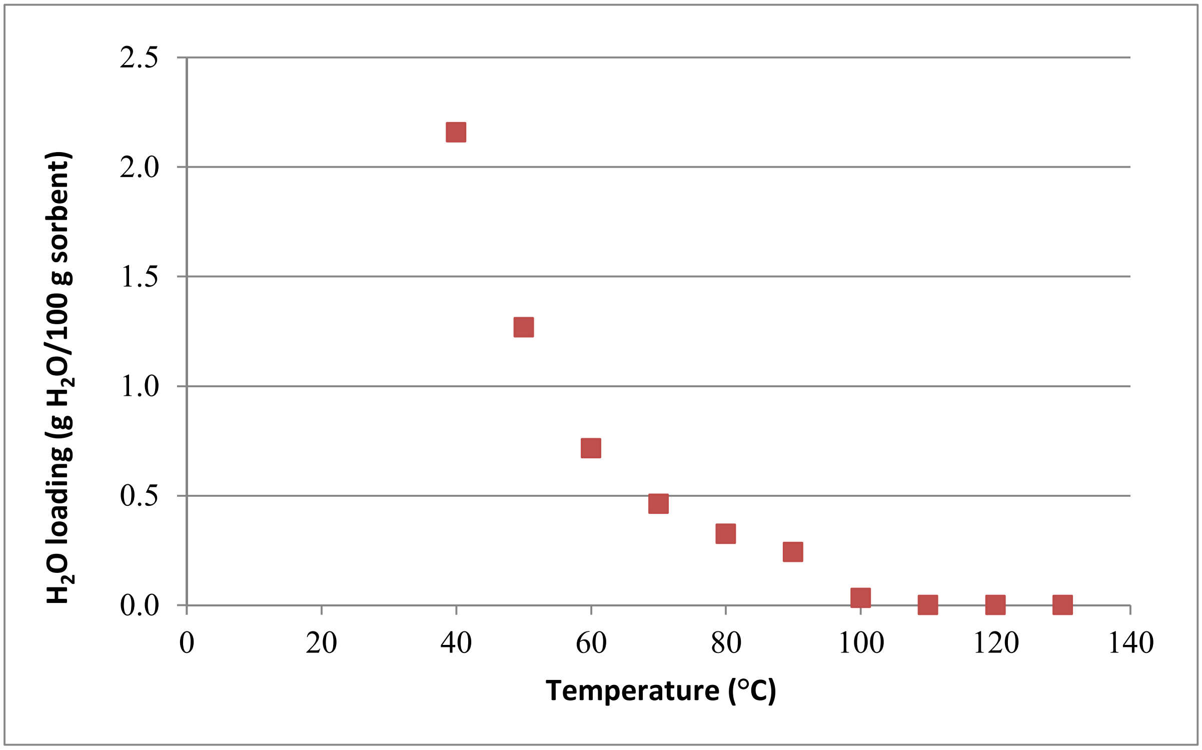


Figure 23. Moisture Uptake on Sorbent BN (N2 with 5.7 vol% H_2O in Gas)

Several tests were conducted to determine the potential impact of oxygen on the adsorption capacity of sorbent BN. The CO_2 loading of BN following pretreatment with either 120°C air, or a 120°C gas mixture containing 6% oxygen, 15% CO_2 and N_2 amine oxidation at elevated temperatures is compared to BN without pretreatment in Figure 24. The data suggests a significant detrimental impact on BN after exposure to 6% oxygen or air at elevated temperatures. This is likely due to the oxidation of the amines. This is a concern in the ADAsorb system because CO_2-lean flue gas was intended for use as the carrier gas between the regenerator and the adsorber. Flue gas typically contains 3 to 6% oxygen.

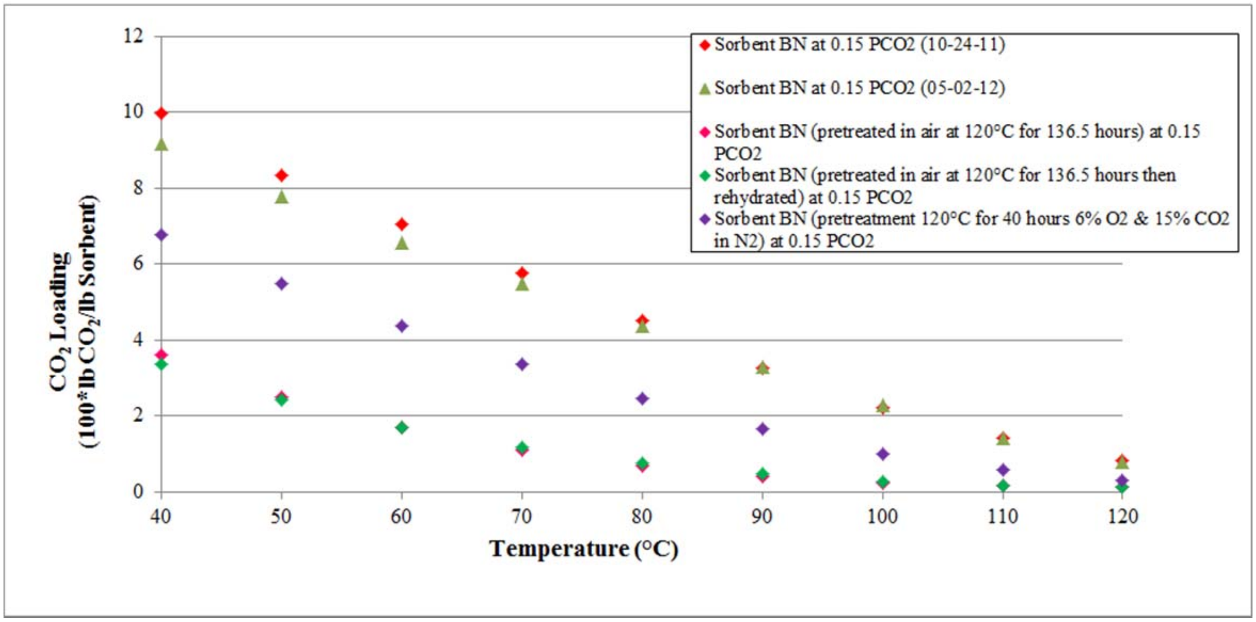


Figure 24. Impact of exposure to oxygen at elevated temperatures on BN performance

3.6.5 Heat of Reaction

Several different experiments were requested in order to assess the heat of adsorption of CO_2 as well as the heat of adsorption of moisture. The tests were challenging and several results were discarded due to experimental uncertainties. Two tests results exhibited sufficient agreement between multiple runs and provided some value. The tests were run in a standalone differential scanning calorimeter (DSC) system (no TGA), so the actual uptake of CO_2 was not measured. Although the heat generation was measured directly by the DSC, the CO_2 uptake was estimated from isotherms shown previously in Figure 15.

The first test was designed to measure the heat of reaction between the sorbent and CO_2 (no moisture). This experiment was characterized by exposing the sorbent first to dry N_2 at 120°C and then to a dry 15% CO_2/85% N_2 blend at 40°C. Based on separate tests, it was estimated that the drier gas would lead to a 20% reduction in the total CO_2 loading. By estimating the CO_2 loading and using the heat generated (measured by the outside lab) the heat of reaction for the dry gas (in which CO_2 uptake only should be generating heat) was ~77 kJ/mol CO_2.

A second experiment was designed to measure the total heat of reaction, taking into account both the heat of adsorption between the sorbent and CO_2 and the heat of adsorption between the sorbent and

H_2O. This experiment was characterized by exposing the sorbent to dry N_2 at 120°C, allowing the sorbent to reach equilibrium, and then exposing the sorbent to a 15% CO_2/85% N_2 blended gas that was saturated at 40°C. Using the estimated CO_2 uptake under those conditions, the effective heat of reaction (which should include the CO_2 uptake and the H_2O adsorption) was 82 kJ/mol CO_2. It is important to realize that many different assumptions were made to arrive at these estimates for the heat of reaction. Thus, they were treated only as order of magnitude estimates. To fully quantify the heat generated during the adsorption process, pilot testing was required to accurately measure the impacts of CO_2 and H_2O adsorption under actual process conditions.

ADA combined empirical data collected from prior field testing and laboratory experiments and used an iterative modeling approach to enhance the design effort. Results from this effort played an important role in the FEED step of the development process. Results from the modeling effort were used to optimize the 1 MWe design.

In previous sections, laboratory analysis was used to compare different sorbents. Based on these analyses, sorbent BN was selected to be the sorbent used during pilot testing and the techno-economic assessment.

3.6.6 Physical Strength and Attrition

Two different types of tests were completed to assess the physical strength of sorbent BN. First, the material was tested in an “as received” condition, which is identified as “beaded”. The sorbent is currently manufactured as part of another commercial process. The “as received” particle is spherical and larger than appropriate for the ADAsorb process. A beaded batch of sorbent with a mean particle diameter of 635 µm was tested to measure the crush strength. The crush strength test was completed using the beaded material. The crush strength was identified as the amount of weight placed on a bead when an audible crack was observed. This occurred at 2 kg/bead.

In addition to the crush strength test, a research entity was hired to evaluate the attrition properties of the sorbent that was beaded (“as received”) as well as sorbent that had been ground to a finer size for use in the ADAsorb process (approximate mean diameter was 100 µm). The likelihood of attrition was assessed using a jet-cup attrition test.

Two samples, identified as "beads" and "crushed sorbent", were received by the test facility. Figure 25 shows the particle size distribution (PSD) of these materials obtained by using Sympatec Halos III analyzer.

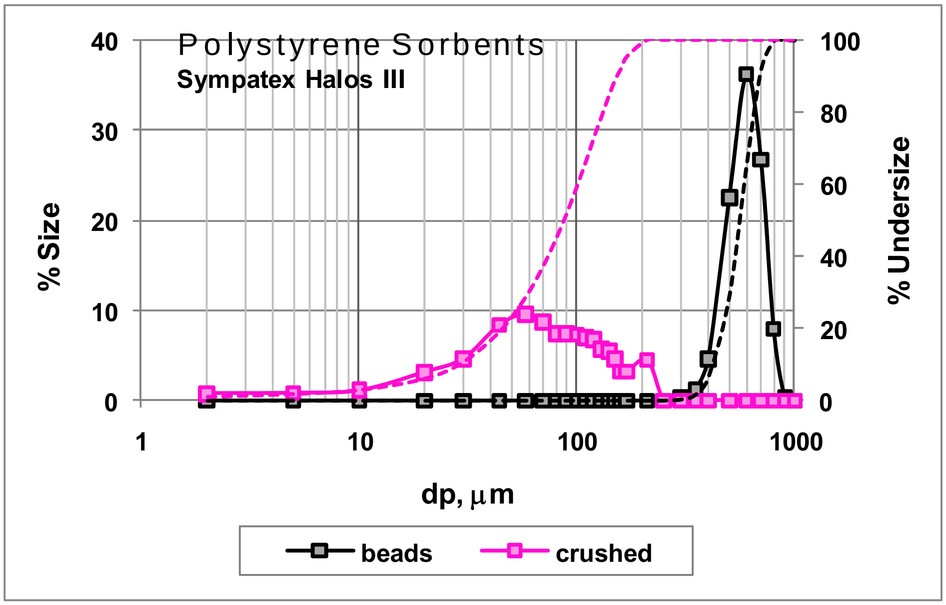


Figure 25. Particle Size Analysis of Two Sorbent Materials.

The particle size distributions for both initial and attrited samples of sorbent beads are shown in Figure 26.

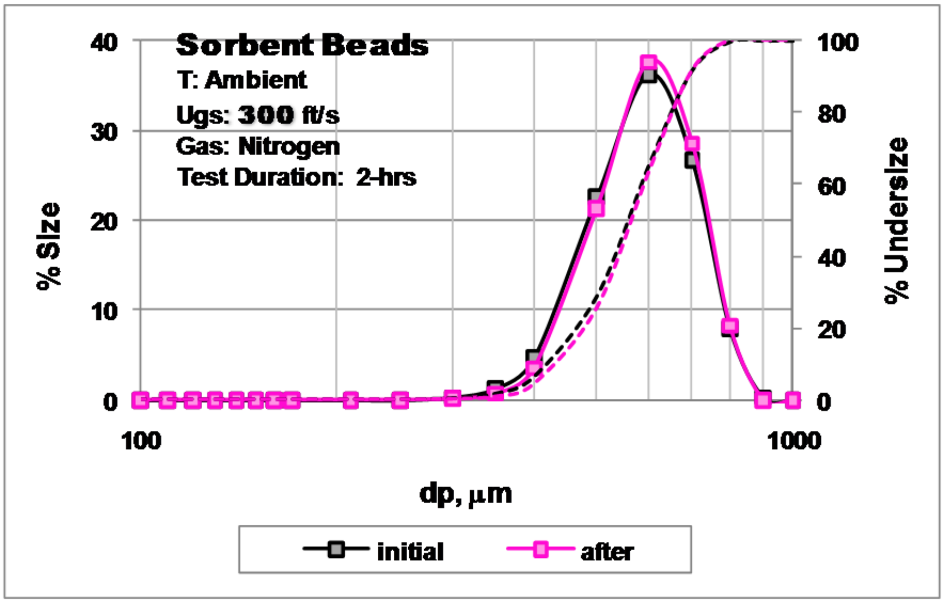


Figure 26. Attrition Characteristics of "Sorbent Beads" at 300 ft/s.

The crushed sorbent material contained almost 20% fines less than 44 microns. Therefore, these fines were selectively removed using sieves. The particle size distributions for both initial and attrited samples of fines-free crushed sorbent is shown in Figure 27.

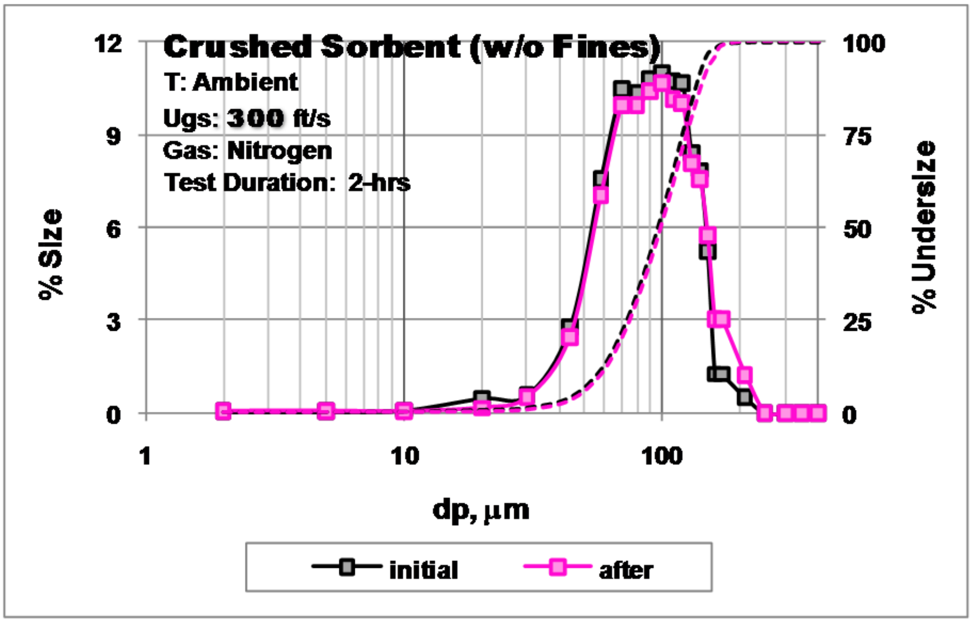
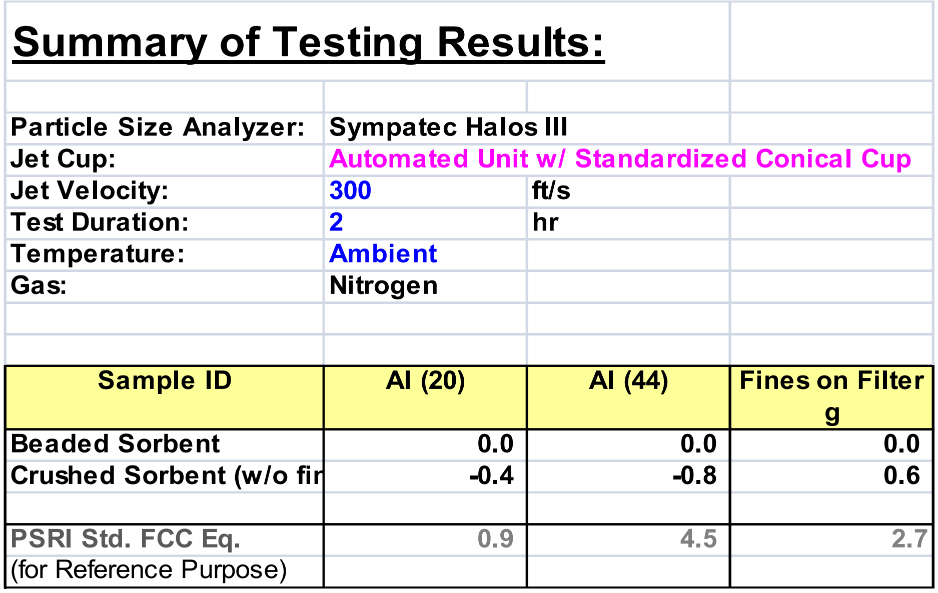


Figure 27. Attrition Characteristics of crushed sorbent at 300 ft/s.

The following test conditions were used in comparing two sorbents and test results are summarized in Table 9.

Table 9. Summary of Attrition Testing Results.



Summary of Testing Results:

Particle Size Analyzer:	**Sympatec Halos III**		
Jet Cup:	**Automated Unit w/ Standardized Conical Cup**		
Jet Velocity:	**300**	**ft/s**	
Test Duration:	**2**	**hr**	
Temperature:	**Ambient**		
Gas:	**Nitrogen**		

Sample ID	AI (20)	AI (44)	Fines on Filter g
Beaded Sorbent	**0.0**	**0.0**	**0.0**
Crushed Sorbent (w/o fir	**-0.4**	**-0.8**	**0.6**
PSRI Std. FCC Eq.	**0.9**	**4.5**	**2.7**
(for Reference Purpose)			

Results based on both attrition indices 20 and 44 (see Table 9) indicated that both the crushed sorbent and beads were highly attrition resistant. The attrition indices are much lower than the reference FCC equilibrium catalyst evaluated previously at the test facility.

The results of the crush strength and attrition testing revealed that this material exhibited promising attrition resistant characteristics. The economic analysis of the overall process, described in other reports, conservatively used the attrition rate of FCC catalysts to estimate the sorbent replacement requirements.

Note that in a commercial application, sorbent BN would be produced at the size required for the ADAsorb process and not ground. The manufacturing process inherently produces spherical particles that tend to plastically deform rather than break. This is ideal for a fluidized bed process and should result in very low attrition. However, due to manufacturing constraints at the quantities required for pilot testing, the material could not be manufactured at the size required. Breaking the resin beads will result in sharp edges and fines that will result in higher attrition than if spherical beads were available for use. Longer term pilot or full-scale testing with appropriately-sized beaded sorbents would be required to quantify actual rates of attrition for this material.

3.6.7 Cold Flow Modeling

As previously mentioned, two cold flow models were used during this program. Initial testing was conducted using a PSRI Inc. standard test setup that provided information on solid handling characteristics and could be used to measure heat transfer in air. The second cold flow model was designed to mimic some of the basic fluidization and conveying processes that were designed into the pilot scale unit and provided information on sorbent fluidization and handling properties in a non-reacting flow environment.

Sorbent characteristics found using the PSRI Inc. cold flow model are given in Table 10 and Table 11.

Table 10. Sorbent BN Properties Derived through Cold Flow Modeling

Geldart Classification	A/B Transition	
Sauter Mean Particle Size	158	µm
90% size Distribution	80	µm
Particle Sphericity	90	unitless
Particle Density	0.71	g/cc
Particle Porosity	0.39	unitless
Bulk Density (Fluffed)	0.38	g/cc
Void Fraction (Fluffed)	0.46	Fraction
Bulk Density (Packed)	0.42	g/cc
Void Fraction (Packed)	0.41	Fraction
Minimum Fluidization Velocity	0.86	ft/s
Particle Aspect Ratio	0.74	unitless
Heat of reaction	82	kJ/mol CO_2

PSRI's initial cold flow modeling was conducted using Sorbent BN with a median particle size (dp,50) of approximately 95 microns. This was deemed satisfactory for the planned nominal size of 100-micron material, which is a Geldart Group A material.

Seven tests were conducted with the 95-micron sorbent, and data were collected to measure bed densities, heat transfer coefficients, fines entrainment rates, CO_2 concentrations, bed pressure drop fluctuations, and bubble properties at the center of the bed. A listing of the test conditions and results appear in Table 11.

The measured bed densities were a strong function of the fluidizing gas velocity in the bed, decreasing linearly with increasing velocity. When CO_2 was added to the circulating gas to a concentration of approximately 6% by volume, the measured bed density increased by roughly 5%, due to the adsorption of CO_2.

The addition of CO_2 had a dramatic effect on the measured solids entrainment rates. At a fluidizing gas velocity of approximately 1.44 ft/s, the entrainment rate with CO_2 present was only 40% of what was measured without CO_2.

Table 11. PSRI Cold Flow Testing of 95 µm Sorbent

Gas Velocity	Bed Density	Heat Transfer Coefficient	Std. Deviation of ΔP	Entrainment Flux	CO_2 Concentration
ft/sec	lb/ft^3	BTU/hr-ft^2-°F	in. wc	lb/s-ft^2	%
0.50	19.21	75.7	0.504	0.00284	< 0.1
1.01	17.09	78.2	0.550	0.03713	< 0.1
1.42	15.33	78.9	0.568	0.09086	< 0.1
0.51	20.10	75.1	0.607	n/a	5.68
1.03	18.06	77.4	0.715	0.00799	6.57
1.46	16.60	76.8	0.693	0.03644	6.23
1.75	15.56	80.7	0.756	0.07048	6.16

Heat transfer coefficients measured both with and without CO_2 in the gas mixture were in the range of 75 to 80 BTU/hr-ft^2-°F (425 to 455 W/m^2-K), increasing slightly with increasing gas velocity. This value is notably higher than values previously calculated using the Molerus Technique (1992 & 1993), which were 45 to 50 BTU/hr-ft^2-°F (255 to 285 W/m^2-K).

The standard deviation of the bed pressure drop measurements was very low, generally less than 1 inch of water column. This represents smaller gas bubbles and good mixing, which result in good solids-gas contacting.

Additional testing by PSRI was conducted to generate data needed to identify the necessary particle size distribution for pilot plant operation. PSRI tested two particle size distributions . The first sample (Fine), with a Sauter mean diameter of 78.7 µm, was material crushed by ADA, with the fines content reduced by elutriation in the PSRI 12" fluidization column. The second sample (Coarse), with a Sauter mean diameter of 255.8 µm, was spherical bead material that PSRI crushed and screened to -40/+120 mesh (125 to 420 µm). Mean particle size is expressed by the Sauter mean diameter, dpSV, which is the diameter of the spherical particle with a surface area to volume ratio of the average size particle, and as such it is the hydrodynamically equivalent particle size.

The estimated size distribution used as the basis for the pilot plant design was developed using the PSRI fluidization regime map. It represented the coarsest size distribution that would still be a Group A powder that could be operated at the highest velocity, while still maintaining a turbulent fluid bed.

The range of PSRI data brackets the pilot plant design basis point, which is labeled as "Pilot Plant Basis" on Figure 28. The fine material exhibited Group A behavior and fluidized very well, but above about 1.5 ft/s, the bed level started to become indistinct, and it appeared to be close to the maximum velocity for this size distribution. This material was clearly too fine for the pilot plant design basis of 4.0 ft/s. The coarse material falls in the Group B range on the fluidization regime map, and PSRI confirmed that it exhibited Group B behavior. They also confirmed that it could have run at well over the highest tested velocity of 4.27 ft/s. This material was too coarse for the pilot plant.

The entrainment rate measured for the fine material was higher than predicted by Technip, while that for the coarse material was lower than predicted as displayed in Table 12. The entrainment fluxes design basis for the three stages in the pilot plant adsorber were between 0.3 and 0.6 lb/s.ft^2.

Table 12. Entrainment Flux

Material: Fine (78 µm)	Entrainment flux (lb/s.ft^2)	
Velocity (ft/s)	Predicted	Predicted
0.5	0.00284	0.001
1.01	0.03713	0.014
1.42	0.09086	0.059
Material: Coarse (255 µm)		
2.38	0.00438	0.023
3.35	0.02623	0.084
4.24	0.08608	0.235

The heat transfer coefficient for the fine material was approximately 80 Btu/lb.ft^2.°F (455 W/m^2-K), and approximately 65 Btu/lb.ft^2.°F (370 W/m^2-K) for the coarse material. The initial pilot plant was designed using a value of 52 Btu/lb.ft^2.°F (295 W/m^2-K), based on lab scale testing. Integrating this information into the 1 MWe pilot process design reduced the cooling coil surface area inside the adsorber vessel, which resulted in a reduction in vessel height. Additionally, plant personnel from the host site measured the temperature of the cooling water source that would be used by the pilot plant. The temperature of the cooling water was 12°C (54°F), which was significantly lower than the temperature used as the design basis. The project team decided to reduce the design cooling water temperature used to calculate the amount of required cooling coil surface area inside the adsorber. Integrating both the lower cooling water temperature and increased heat transfer coefficient into the process design reduced the overall height of the adsorber by approximately 48 ft. The reduction in adsorber height reduced costs in several areas:

- Structural steel in module fabrication.
- Overturning moment on structure, thus reducing complexity of foundation and overall size.
- Fabrication costs of adsorber vessel.
- Piping installation.

Of the two distributions tested, one is too fine and the other too coarse, so the target lies between the two. For the pilot plant, Technip stated that it was more important to have a Group A material, and to reduce the fluid bed density in order to minimize adsorber pressure drop, than to minimize entrainment rate. The exact location of the Group A – Group B boundary is uncertain, and Technip believed it was preferable to err on the finer side to ensure that the target is a Group A material. For this reason, the final recommended sorbent size distribution should be somewhat finer than that used for the pilot plant design. The recommended target particle size distribution is displayed in Table 13 and Figure 28.

Table 13. Recommended Particle Size Distribution

dp, micron	wt% less than
40	0.7
60	1
80	1.8
100	3.6
120	7.6
140	15
170	31
200	55
220	75
250	88
270	94.5
300	98
320	99.5
Sauter mean	169.1 µm

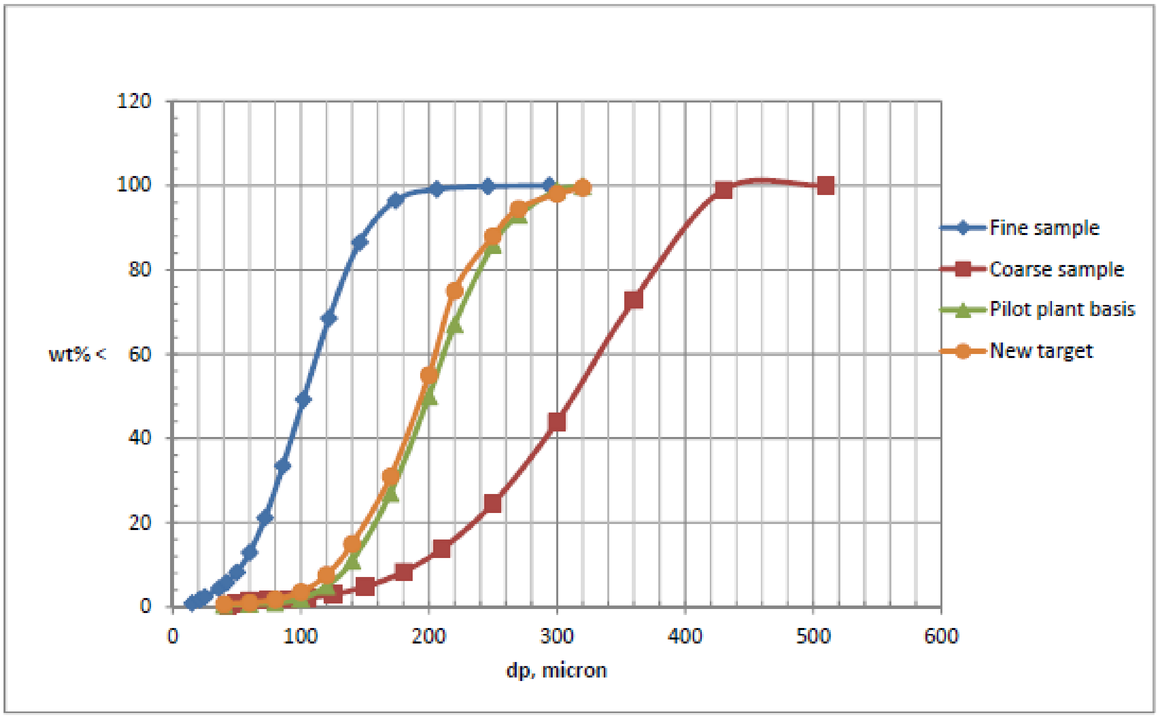


Figure 28. Particle Size Distribution

The second model was constructed with two stages and multiple cyclones to allow investigations into fluidization regimes expected in the pilot facility. A photo of the model is shown in Figure 29. The benefit of this model was that it allowed for systematic testing of sorbent material at a wide range of fluidization gas velocities, and was used for operator training. Additionally it allowed for testing of low pressure drop gas distribution plates which were ultimately used during pilot testing.

Figure 29. ADA Cold Flow Model

Because the model was equipped with two-stage cyclones, downcomers, and diglegs, it was possible to run the cold flow model at high velocities that met and exceeded the design velocities in the pilot and to verify functionality under cold flow conditions. During the cold flow model parametric testing, observations could be made at the onset of bubbling fluidization, turbulent fluidization, and circulating fluidization as entrained sorbent was carried into the cyclones and then returned into the beds.

During parametric testing of the sorbent within the model, multiple sorbent samples were taken at different velocities. Ultimately, the sorbent was fluidized extensively at the pilot designed gas velocity and fine particles that were not collected by the cyclones were collected in a baghouse and removed from the system. This allowed bed samples of conditioned sorbent to be taken and analyzed for a particle sample size. These samples formed the basis of the target particle size distribution for sorbent production and pilot operations.

3.6.8 Summary of Sorbent Physical and Thermal Properties

As discussed above, several different physical properties were measured by outside laboratories as well as by ADA. The results are summarized in the following bullet list:

- Pore volume: 0.27 mL/g (provided by manufacturer)
- Particle density: 0.646 g/cm^3 – As shipped from the manufacturer – measured at Adsorption Research, Inc.
- Density
  - Bulk density: 0.586 g/mL (36.5 lb/ft^3) – After being treated with saturated N_2 at 55°C – measured at Adsorption Research, Inc.
  - Bulk density: 0.445 g/mL (27.8 lb/ft^3) after treatment at 40C in 4% H_2O in N_2 – measured by ADA
  - Tap density: 0.458 g/mL (28.592 lb/ft^3) after treatment at 40C in 4% H_2O in N_2 – measured by ADA
- Sorbent heat capacity: 1.05 kJ/kg·K – measured by FAI laboratories
- Sorbent thermal conductivity: 0.08 W/m/K (for polystyrene, which is the main substrate component)
- Crush strength: 2 kg/bead, Measured by Adsorption Research, Inc.

3.6.9 Sorbent Cost

While the sorbent cost is considered a key sorbent property, it cannot be assessed using laboratory experiments. At the onset of the current program, a goal sorbent cost of $5/lb was assumed. Although a commercial CO_2 capture sorbent cost is not yet known, the attrition results and subsequent economic sensitivity analyses suggest that it may be possible for a commercial sorbent to be more expensive than $5/lb without making the overall process cost prohibitive.

3.7 Process Calculations and Computational Modeling

During design of the pilot system, Technip utilized proprietary models to design the three-stage adsorber and single-stage regenerator. Key sorbent characteristics such as particle density, particle size, and thermal conductivity were inputs into the model.

ADA also worked with NETL and the Carbon Capture Simulation Initiative (CCSI) program to model the adsorption and regeneration process. The CCSI modeling was conducted using a one-dimensional computational model for a bubbling fluidized bed reactor developed at NETL. ADA provided inputs for the model. The CCSI team modeled reaction kinetics for the sorbent using a lumped parameter equilibrium and kinetic model developed at NETL and fitted to experimental data obtained using TGA data provided by ADA.

The CCSI model predicted an overall removal of CO_2 from the process gas stream of 58.07% and an achieved working load of 4.5 g CO_2 /100 g sorbent, compared to the design condition of 7 g CO_2/100 g sorbent. The CCSI team identified two possible reasons for the poor modeled performance of this process. The first of these is due to the predicted non-isothermal behavior of the regenerator in this process. The CCSI model highlighted the strong cooling effect as sorbents loaded with CO_2 were introduced at the top of the regenerator. Based on the model, the temperature of the solids leaving the regenerator was only 104°C. The addition of cool solids immediately decreases the local temperature of the bed, which is compounded by the rapid release of adsorbed species from the heavily loaded sorbent via endothermic reactions.

The second potential reason for the low working CO_2 capacity from the model is due to the model input data with significantly slower kinetics of adsorption of CO_2 at low temperatures which limited the modeled performance of the adsorber. The CCSI team was not provided with all of the kinetic studies conducted by ADA. Some of the initial results, which were provided to the CCSI team, were conducted with a standard stepwise temperature ramp and the results may have been representative of actual process conditions in the pilot. The model predicted that the uptake of CO_2 appears to be approaching to 85% of the equilibrium loading. This would suggest that the performance of the adsorber could be improved slightly with faster kinetics or longer residence times, but that improvement might be uneconomical.

The regenerator design in the pilot includes features from existing fluidized beds used in the chemical and process industries. A bank of heat transfer tubes is located within the bed of sorbent that takes up 32% of the cross-section of the regenerator. There are a few inches of sorbent bed above and below the bank of heat transfer tubes. If the fluidization velocity is too low, the sorbent will bubble rather than fully fluidize in the spaces above and below the tubes. Since the steel tubes in the regenerator significantly restrict the cross sectional area for fluidization flow, the velocity within the tube zone became significantly greater than the velocity above and below the tubes.

One of the recommendations made by CCSI to improve performance was to use a bottom sorbent exit on the regenerator. The high velocity within the tube zone prevents sorbent circulation to the regenerator bottom, making the bottom exit design infeasible. A bottom exit would be starved of sorbent flow when the tube zone fluidization velocity exceeded the sorbent transport velocity.

Another potential option would be to operate the regenerator at a lower velocity, resulting in bubbling beds above and below the heat transfer tubes, and fluidized flow within the tube zone. During bubbling operation, the sorbent would be fluidized within the tube zone, but the residence time associated with fluidization would be a small fraction of the overall residence time in the regenerator. The majority of the sorbent residence time would occur in the higher density, bubbling bed sections above and below the tubes, where the contact with fluidization gas and heat transfer occur much slower.

A baffle was added at the sorbent inlet of the regenerator to limit potential impacts of cold sorbent entering the bed by directing the new sorbent to a small area on the top of the bed. Regenerator temperature measurements made during pilot testing indicated that the non-isothermal operation predicted by the CCSI team did not exist during pilot operation at the sorbent recirculation rates tested in the pilot, which were 30% to 60% below design. It is possible that as the sorbent recirculation rate approaches the design rate, the behavior modeled by the CCSI team may become more of a concern.

4. Process Design

4.1 Conceptual Design - 550 MW

A full-scale conceptual design was developed using the fundamental process technology identified during the initial design activities, and characteristics of the selected sorbent. Design activities were conducted during Task 2 of the project through the series of full-scale conceptual design activities, as described by the subtasks listed below.

- 2.3: Preliminary Conceptual Design: 550 MW
- 2.4: Gather Additional Design Data
- 2.5: Review Results of Process Model and Sorbent Selection
- 2.6: Refine Commercial Scale (nominal 550MW) Conceptual Design

The specific subtask activities have been combined for clarity in this report, but were conducted iteratively during project execution.

The solid sorbent-based CO_2 capture process developed during this project, referred to in this report as the ADAsorb Process, employs the use of a dry sorbent characterized by amine functionalization. Several assumptions were made during the development of the preliminary conceptual design based on public literature and laboratory testing, including the following.

- Particles physically resemble polystyrene beads with respect to size
- Particle density: 36.6 lb/ft^3
- Heat of reaction is 587 Btu/lb CO_2 adsorbed
- Adsorption temperature is 40°C
- Regeneration temperature is 120°C
- Flue gas temperature after the SO_2 polishing unit is 57°C

The ADAsorb process integrated into the PC plant is shown in Figure 29. Similar to the solvent MEA process, the ADAsorb process is implemented immediately upstream of the power plant stack.

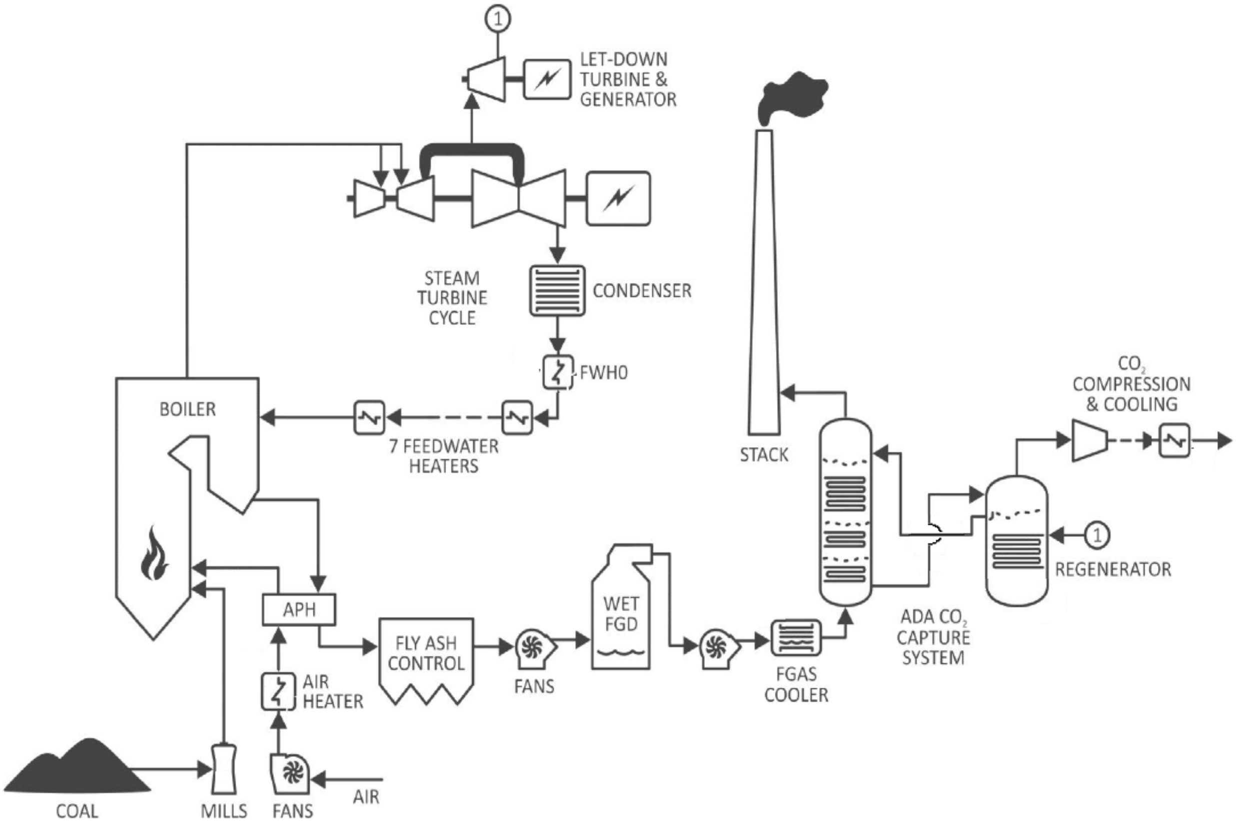


Figure 30. Process Flow Diagram of ADAsorb System Integrated into a Power Plant.

A sketch of the ADAsorb Process is shown in Figure 31. Flue gas is routed from the existing plant through ductwork to the CO_2 capture facility. The amine functionalization on the sorbent can react to form heat-stable salts similar to aqueous amines. Therefore, the SO_2 concentration in the flue gas before entering the CO_2 capture facility must be reduced to levels comparable to those observed in the Econamine process via a similar SO_2 polishing unit. It should be noted that many solids, including the solid evaluated during this project, can be regenerated to recover the functionality of the amines following reaction with SO_2. Thus, as an alternative to a polishing SO_2 scrubber is the ability to remove a bleed stream of sorbent for chemical regeneration.

A flue gas blower is necessary to provide sufficient pressure to pass the flue gas through the CO_2 capture equipment, and return it via ductwork back to the stack. The flue gas is then directed through a cooler, where the temperature is decreased to 40°C. Water vapor is separated from the flue gas in a knockout (KO) drum.

After the KO drum, the cooled flue gas enters a three-stage fluidized bed adsorber. Several adsorber designs were considered, as discussed in Section 3. A fluidized bed TSA process was ultimately selected, based upon process considerations and sorbent characteristics.

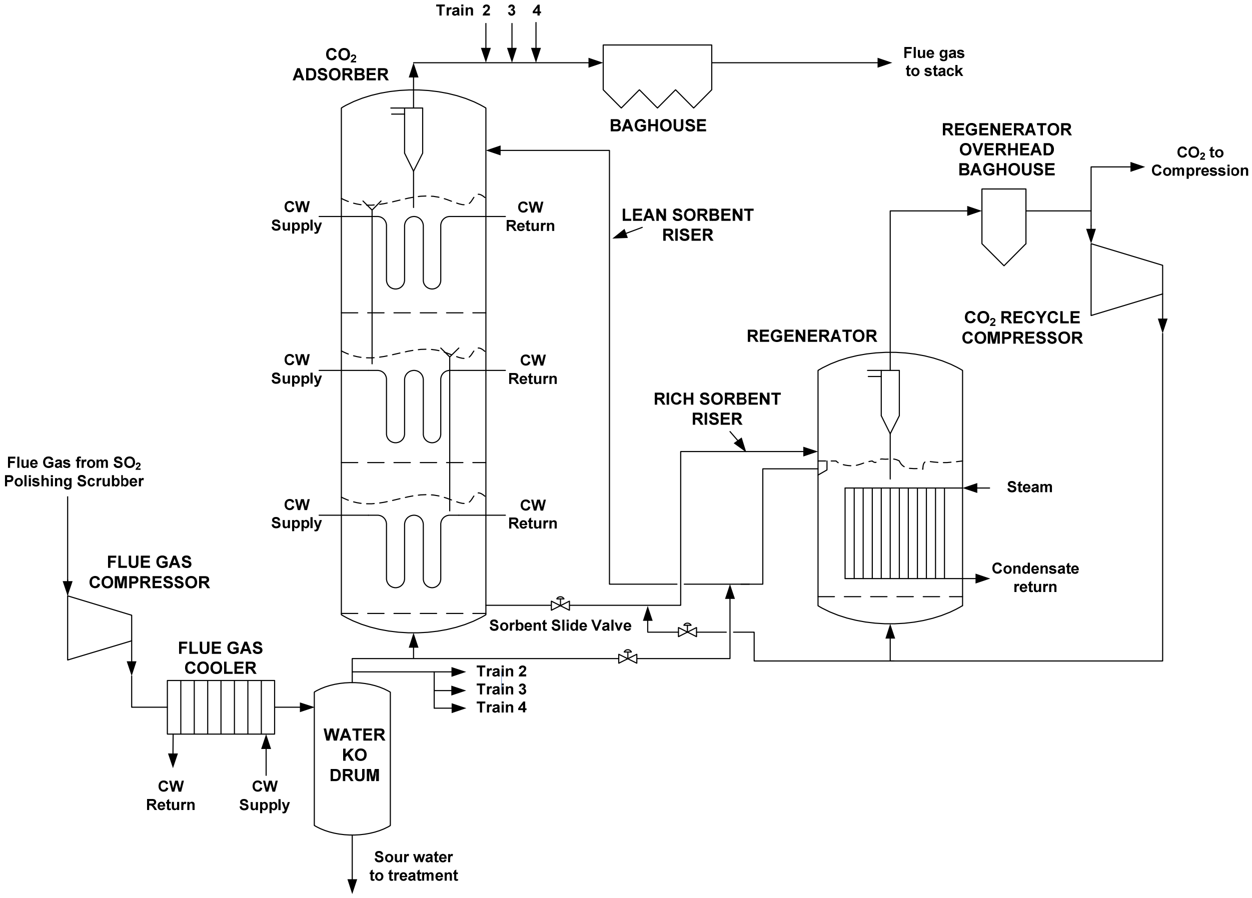


Figure 31. Process Flow Diagram of the ADAsorb Capture Process.

Two-stage fluidized beds are common in the petrochemical industry. Based on provided sorbent properties obtained through laboratory sorbent characterization, results from Technip's proprietary fluidized bed process models, as well as their experience with reactor scale-up, Technip recommended a three-stage fluidized bed as the process technology for the adsorber to approach counter-current flow between the solids and the process gas and achieve the 90% CO_2 removal goals established for the project.

Table 14 indicates the expected pressure and CO_2 loading for the adsorber predicted by laboratory isotherm data, as presented earlier in Figure 15, and output from Technip's proprietary fluidized bed models, where the sorbent loading is the expected equilibrium CO_2 loading. Note that, because of the pressure conditions predicted by the Technip process model, shown in

Table 14 below, the CO_2 working capacity, or difference between the inlet and outlet conditions, for the sorbent in the ADAsorb system was projected to be 10.47 - 3.45 = 7.02% CO_2/sorbent.

Table 14. Predicted CO_2 Partial Pressure and Sorbent Loading in the Adsorber

Adsorber Location		Sorbent Loading (wt% CO_2/sorbent)	CO_2 Partial Pressure (bar)
Stage 3 (Top)	Top	3.45%	0.019
	Bottom	5.79%	0.083
Stage 2 (Mid)	Top	5.79%	0.083
	Bottom	8.13%	0.148
Stage 1 (Bottom)	Top	8.13%	0.148
	Bottom	10.47%	0.214

Each stage in the adsorber is equipped with a bundle of serpentine cooling tubes to maintain isothermal operation. The top bed of the adsorber is responsible for cooling the sorbent from the regeneration temperature (120°C) to the adsorption temperature (40°C), removing the heat generated from the exothermic reaction between CO_2 and the sorbent, and removing the exothermic heat due to the adsorption of the remaining moisture in the flue gas. The cooling coils in the bottom two fluidized beds are responsible only for removing the heat of reaction between CO_2 and the sorbent and the heat adsorption of moisture. Due to this added sensible heat requirement, the top bed depth is larger than the bottom two.

The stages are separated by horizontal steel sections, which allow the flue gas entering at the bottom to flow through the adsorber and fluidize the sorbent. The treated flue gas is released at the adsorber top and routed to a series of cyclones followed by a baghouse, which collects any entrained sorbent.

The staged fluidized bed was selected for several key performance characteristics, including:

- Maximizing the driving force for mass transfer (i.e. CO_2 uptake) by approaching counter-current gas/solids contacting;
- Isothermal operation;
- Operation of fluidized bed reactors has been successfully demonstrated in other industries.

Standpipes function as passages for sorbent to cascade down from one fluidized bed to the next (top to the bottom), while adsorbing CO_2 from the flue gas flowing upward. As the CO2-rich sorbent is withdrawn at the bottom, lean sorbent is simultaneously added at the top of the adsorber. The amount of the rich sorbent discharged is controlled by a sorbent slide valve. The CO2-rich sorbent is pneumatically conveyed with CO_2 to the regenerator via a rich sorbent riser.

In an isothermal (120°C) single-stage fluidized bed regenerator, the rich sorbent releases CO_2 by means indirect steam in an internal heat exchanger. Since the temperature of steam (152°C) extracted from IP/LP crossover section of the PC plant steam turbine is hotter than the regeneration temperature of the solid sorbent (120°C), the expansion of the steam through a back pressure (BP) turbine before entering the exchanger is required to prevent the thermal degradation of the amines impregnated in the solid sorbent and also to generate additional electricity. Saturated steam at 45 psia is obtained at the BP turbine exit where the temperature and pressure of the steam will adjusted accordingly to ensure the amine functional groups are not damaged during the regeneration process.

The higher temperature in the regenerator reverses the reaction between the amine functional groups and the CO_2, thus releasing the CO_2 into the gas phase. The released CO_2 exits the top of the regenerator and is directed through a baghouse. A portion of the CO_2 is routed to the CO_2 recycle blower, while the remaining CO_2 is sent to the compression plant similar to the MEA process. The CO_2 recycle blower provides enough pressure for the CO_2 to promote adequate fluidization inside the regenerator and to convey sorbent from the adsorber discharge to the regenerator in the rich sorbent riser. To complete the cycle, the CO_2-lean sorbent from the regenerator is transferred via a sorbent riser, using the conveying media of CO_2-lean flue gas. CO_2-lean flue gas is utilized to prevent adsorption of CO_2 by the regenerated sorbent prior to returning the sorbent to the adsorber. Using the assumptions from the preliminary conceptual design period, four adsorption/regeneration trains were estimated to process all the flue gas being exhausted from a 550 MW_{net} subcritical coal fired power plant.

The solid sorbent chosen for pilot testing collects CO_2 in an exothermic reaction resulting from chemisorption of CO_2 onto an amine covalently bonded to the sorbent substrate. The sorbent releases CO_2 at elevated temperatures. To prevent heat generated in the reaction from heating the sorbent and quenching further adsorption, cooling coils within the fluidized beds remove generated heat and maintain isothermal conditions throughout the adsorber.

One of the design considerations during the conceptual design period was sorbent conveying within the system. One of the concepts considered was bucket conveyors, which are commonly used to move large volumes of solids. Long-term maintenance costs associated with the required moving parts resulted in a decision to use dense-phase conveying. Laboratory results indicated that at elevated temperatures during regeneration, the presence of oxygen in flue gas may oxidize and degrade functional amines on the sorbent. For this reason, CO_2 product gas was specified as a more suitable option to reduce risk of premature sorbent degradation. Therefore, in the conceptual design, the CO_2 product stream gas to convey hot sorbent from the regenerator to the adsorber.

The sensible heat advantage of solid sorbents compared to solvents is offset by the ease of sensible heat recovery in a cross heat exchanger with an MEA system that uses heat from regenerated solvent to preheat CO_2 laden solvent before it is regenerated. This is due to the inherent ease with which liquids can be pumped through efficiently designed heat exchangers and the physical difficulties of exchanging sensible heat of solid particles while conveying sorbent particles between the adsorber and regenerator. A cross heat exchanger was not designed for this project, but was considered as part of another DOE project, DE-FE0012914, and initial findings will be presented in Section 4.2.

Flue gas from the outlet of the wet scrubber (FGD) equipment flows into the CO_2 scrubber. Following CO_2 removal, the flue gas flows through particulate control devices to separate any sorbent that has been carried out of the fluidized bed. Larger particles will be returned to the adsorber and fine particles will be removed from the system. The low CO_2 flue gas will then be directed to the main stack. Steam for regeneration heat is extracted from the crossover of the IP and LP turbines before being expanded to 166°C and 85 psig (5.9 bar) through a backpressure turbine.

To treat the flue gas from a large (i.e. ≥ 500 MW) coal-fired power plant, four trains of the process shown in Figure 31 would be necessary based on the properties of one specific sorbent. A plot plan of the integration of the process into the coal-fired power plant is provided in Figure 32.

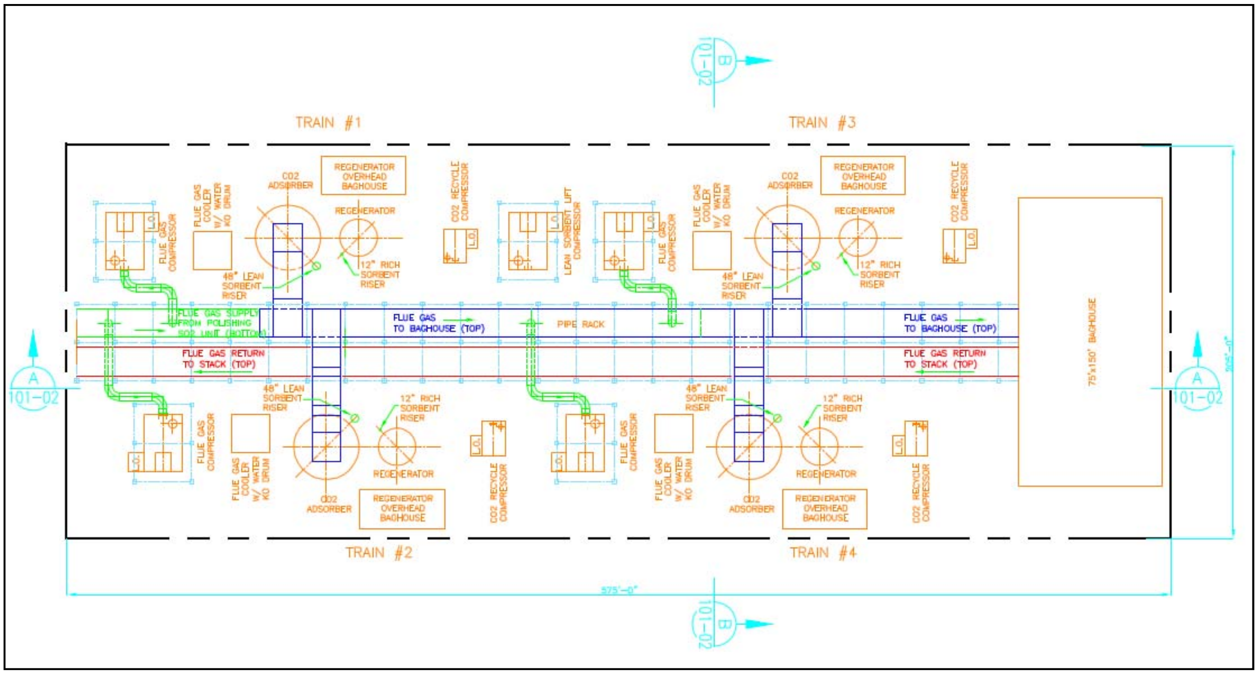


Figure 32. Plot Plan of the ADAsorb system at a Full-Scale Coal-Fired Power Plant

4.1.1 EPC Review of Commercial Scale Process Economics

ADA subcontracted Stantec Consulting to provide a techno-economic assessment which compared the basic ADA solid sorbent system design to a conventional MEA system. During the assessment, the effort indicated the relative cost contributions of capital equipment, operating costs, CO_2 compression costs, and fuel costs using an Illinois #6 bituminous coal as the baseline fuel. Costs of the initial assessment were determined to be slightly less than the costs associated with an MEA system, but well within the error range of such estimates.

The TEA was able to break down the costs associated with electricity generation with the solid sorbent post-combustion capture system. The biggest drivers of costs associated with post-combustion capture were with capital equipment, sorbent cost and attrition, and energy penalty. In order to reduce costs, the most effective method is to improve sorbent characteristics so that less capital equipment is necessary and the overall power plant size can be reduced due to a reduction in energy penalty. In order to accomplish these goals, a sorbent must have the highest working capacity possible, the fastest kinetics possible to reduce reaction vessel size, and be regenerated at the lowest temperature possible to reduce consumption of high-value steam.

Preliminary cost estimates for the ADAsorb Process demonstrated a lower cost of electricity impact when compared to the established MEA process. Vendor quotes and other historical database information were used to calculate the incremental COE increase. As determined using the Association for the Advancement of Cost Engineering International (AACE) methodology, the level of cost accuracy for staged fluidized bed process design is defined by a Class IV/V estimate, which is approximately in a range of ±50 percent accuracy.

Preliminary estimates indicated a PC plant featuring the ADAsorb Process could lead to improved economic viability as well as a greater technical performance than that of the MEA process published in the DOE Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Rev. 2.23. The COE is lowered by three percent and the net plant efficiency is increased by 13 percent for the initial study. Compared with the PC plant without carbon capture (Case 9) published in the DOE report, the PC plant featuring ADA's solid sorbent technology results in a COE increase by 80 percent from 67.6 (escalated to 2015 dollars) to 121.8 mills/kWh, which is greater than the DOE target of 35 percent increase in COE. This gap can be reduced by improving sorbent performance (working capacity, attrition resistance and regeneration energy), which is the largest driver of system performance. Costs could also be reduced by optimization of the steam condition at IP/LP crossover to eliminate a need for the BP turbine and to integrate into the steam cycle in a more efficient manner. Additionally, it may be possible to include heat integration either within the CO_2 capture process or between the CO_2 capture process and the power plant. Finally, the most critical element for cost reduction is to reduce size of capital equipment or the number of vessels due to increases in sorbent performance, which will reduce the capital cost component, which is the largest contributor to COE increase.

A summary of the techno-economic assessment conducted in the second quarter of 2011 is shown in Table 15. All costs are estimated in 2015 US dollars.

Table 15. Technology Comparison Results for Solid Sorbent and MEA.

Description	Unit	Econamine1 BEC as per DOE	Sorbent1 BEC as per DOE
Gross Power Output	kW_e	672,700	656,125
Electrical Output – BP Turbine	kW_e	0	47,369
Auxiliary Load Summary	kW_e	122,740	153,534
Net Power Output	kW_e	549,960	549,960
Thermal Input	kW_{th}	2,102,644	1,865,320
Coal Flow Rate	lb/hr	614,994	545,581
Net Plant HHV Efficiency	%	26.2	29.5
CO_2 Captured	lb/hr	1,313,960	1,165,561
Raw Water Consumption	1000 gal/D	8,081	6,460
Regen. Energy	Btu/lb CO_2	1,530	1,139*
Bare Erected Cost	2015$	1,415,831,000	1,294,480,737
COE (2015$)	**mills/kWh**	**125.0**	**121.8**

* The regeneration energy reported does not include any heat integration or optimization strategies.

The modeling results of the supported amine sorbent highlight that the activated carbon sorbents would not be cost-competitive. Results of sorbent testing indicated that the working capacity of the activated carbon sorbents was less than half of the working capacity of the supported amine sorbent BN. As a result, the capital costs associated with the reduced working capacity would increase significantly because larger adsorption vessels would be needed to accommodate the greater amount of sorbent necessary to facilitate 90% CO_2 capture from the flue gas. Since reactor vessels are capital equipment and capital equipment was determined to be the most significant cost driver, it can be assumed that use of activated carbon sorbents in a fluidized bed process would be cost prohibitive.

Various equipment vendors were consulted to analyze the capital costs associated with a full-scale system. In some cases, the laboratory screening data was insufficient to answer all pertinent questions. For example, fixed bed testing results were not sufficient to provide precise reaction kinetic rates for a counter-current system. Since the capital costs of different systems are highly dependent on the required gas/solid contact time, which is dictated by reaction kinetics at the appropriate conditions, results from TGA testing in conjunction with results from the trickle-down reactor were used to extrapolate the required contact time for specific CO_2 sorbents. Although a trickle-down reactor will result in different mass and heat transfer characteristics than a three-stage fluidized bed, it represents a more “ideal” counter-current design and results can help inform design decisions for a fluidized bed system. Results from the trickle-down tests were presented in Section 3.6.3, where laboratory testing suggested that the reaction kinetics under 40°C isothermal conditions and near ambient pressures experienced in the trickle-down reactor were fairly fast, achieving 90% CO_2 capture within seconds at

CO_2 loading of 70 to 80% of equilibrium. The TGA results suggested that adsorption kinetics may actually be much slower as the sorbent becomes saturated with CO_2, requiring tens of minutes to reach equilibrium. The adsorber design conditions for the ADAsorb system to assure 90% CO_2 capture were set primarily on the heat transfer characteristics of the sorbent and the size and time required to maintain isothermal conditions. Furthermore, the ADAsorb adsorption section was designed with flexibility to allow some control over residence time to better evaluate the actual residence time in the adsorber that was required to achieve the CO_2 capture goal.

Results of a techno-economic assessment completed after pilot testing, and insights into potential reductions in energy penalty resulting from heat integration, are included in Section 7.

4.2 Pilot Design – 1 MWe

The detailed design of the ADAsorb pilot facility was completed by Technip with support from ADA. Pilot design details were established based upon laboratory results and characteristics of a single sorbent, identified as sorbent "BN" herein, and in a previous topical report.[34] Additional design parameters included adsorption at 40°C (104° F) and regeneration at 120° C (248° F). The pilot was designed for an optimal sorbent circulation rate of 32,600 lb/hr to achieve a sorbent CO_2 working capacity of approximately 7 g CO_2 / 100 g sorbent, based upon results from laboratory testing.

The detailed pilot design included the necessary P&IDs, a process flow diagram, heat and mass balance sheets, equipment design, general instrumentation, and equipment layout. ADA worked with its subcontractor and equipment vendors to develop a control scheme, gas concentration measurement instrumentation, and performance measurements. These efforts were jointly used to create the necessary documentation packages for the construction company to fabricate and install the pilot equipment used for the 1 MWe pilot plant. A simplified schematic of the system is presented in Figure 33. The key parameters used for the pilot design are presented in Table 16.

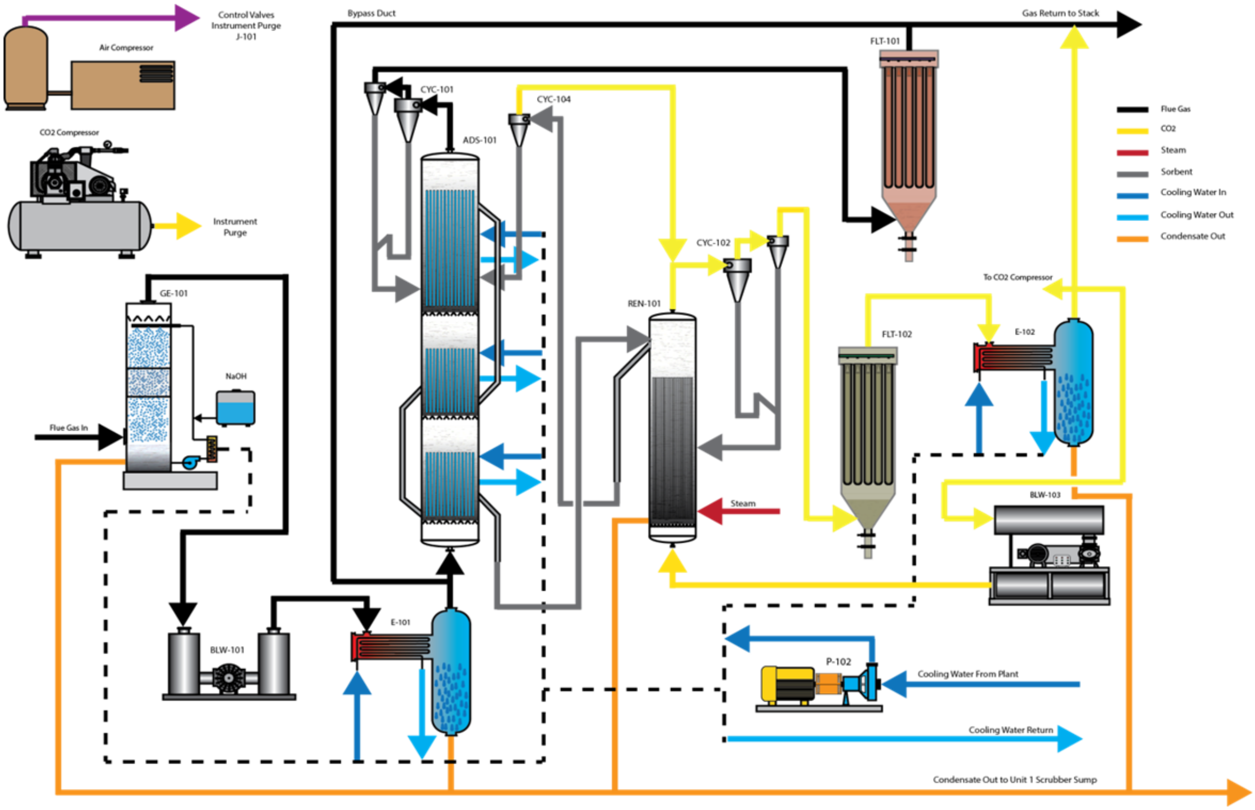


Figure 33. Process Diagram Illustrating Main Vessels Used in the ADAsorb Process.

Table 16. Pilot Design Basis.

Parameter	Location	Value	Units
Flue Gas Flow Rate		1808	acfm
Gas Velocity in Adsorber		4.2	ft/sec
Sorbent Circulation Rate		31107	lb/hr
% CO_2 Captured		90.00%	Weight %
CO_2 Working Capacity		7.00%	gCO_2/g Sorbent
H_2O Working Capacity		1.00%	gH_2O/g Sorbent
CO_2	Fluidizing	343	acfm
	Transfer to REG	67	acfm
	Transfer to ADS	252	acfm
	Total used in Process	662	acfm
	Product	375	acfm
	CO2 Removed within	2172	lbs/hr
Temp	Bottom ADS bed	104	F
	Middle ADS bed	104	F
	Top ADS bed	104	F
	Regenerator	248	F
Adsorber In	CO_2	13.1	%
	O_2	5.65	%
	H_2O	4.85	%
Adsorber Out	CO_2	1.58	%
	O_2	6.73	%
	H_2O	0.78	%
Regenerator Out	$CO_{2\,(dry)}$	93.4	$\%_{(dry)}$
	O_2	0.38	%
	H_2O	7.2	%

4.2.1 Polishing Scrubber

The pilot was designed to treat a slipstream of 3,109 SCFM of flue gas extracted from the host plant downstream of the flue gas desulfurization (FGD) unit at a temperature averaging 57°C (135°F). After leaving the host site, the flue gas is passes through a polishing scrubber, GE-101, that reduces sulfur dioxide (SO_2) concentrations in the incoming flue gas to the design specification of less than 1 ppmv. This scrubber is necessary to reduce the SO_2 levels in the flue gas to minimize the formation of heat-stable salts that would create a loss of amine functionality. The temperature of the flue gas increases within the scrubber due to the heat of reaction of the SO_2 with the scrubber's 20% caustic solution. A built-in slurry cooler utilizing a plate and frame heat exchanger and approximately 125 gpm of water, reduces the gas to a scrubber exit temperature of 40°C (104°F). All pipework upstream and immediately downstream of the scrubber, as well as the scrubber itself, is constructed of fiber reinforced plastic (FRP) to eliminate corrosive effects. A sketch of the scrubber skid is shown in Figure 34.

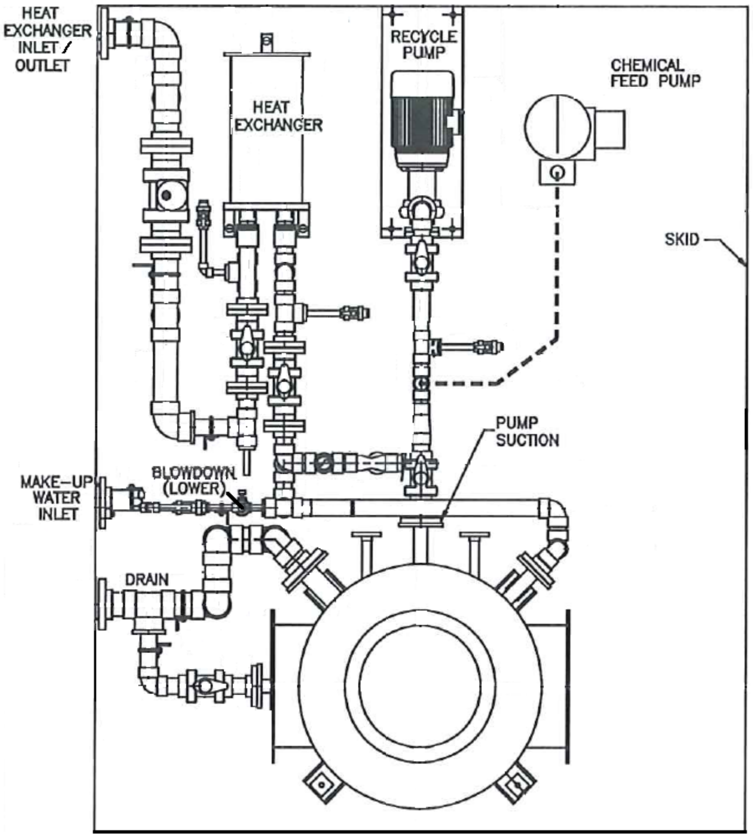


Figure 34. SO_2 Scrubber Skid

4.2.2 Blower and Flue Gas Cooler

A blower, BLW-101, is installed after the scrubber to provide sufficient pressure to overcome the pressure drop of the capture system and allow flue gas to pass through the carbon capture equipment and return it via ductwork back to the original stack. The pilot was designed to operate with a blower back pressure and differential pressure of 7.3 psig and 7.7 psig respectively. This blower is operationally limited by the differential pressure across the blower and is rated for a maximum continuous operational differential pressure of 15 psig. As a failsafe to protect the FRP scrubber and its inlet and outlet piping in the event of ductwork pluggage, a vacuum relief valve is installed between the scrubber and the blower. The pressure drop of the system is due in part to the bed density, bed depth, pressure drop across the distribution plates at each stage, and pressure drops through the cyclones and bag houses. As a result of the flue gas compression, the flue gas temperature is increased.

As the flue gas is compressed within the blower the temperature increases by approximately 50°C in accordance with the blower curve provided in Figure 35.

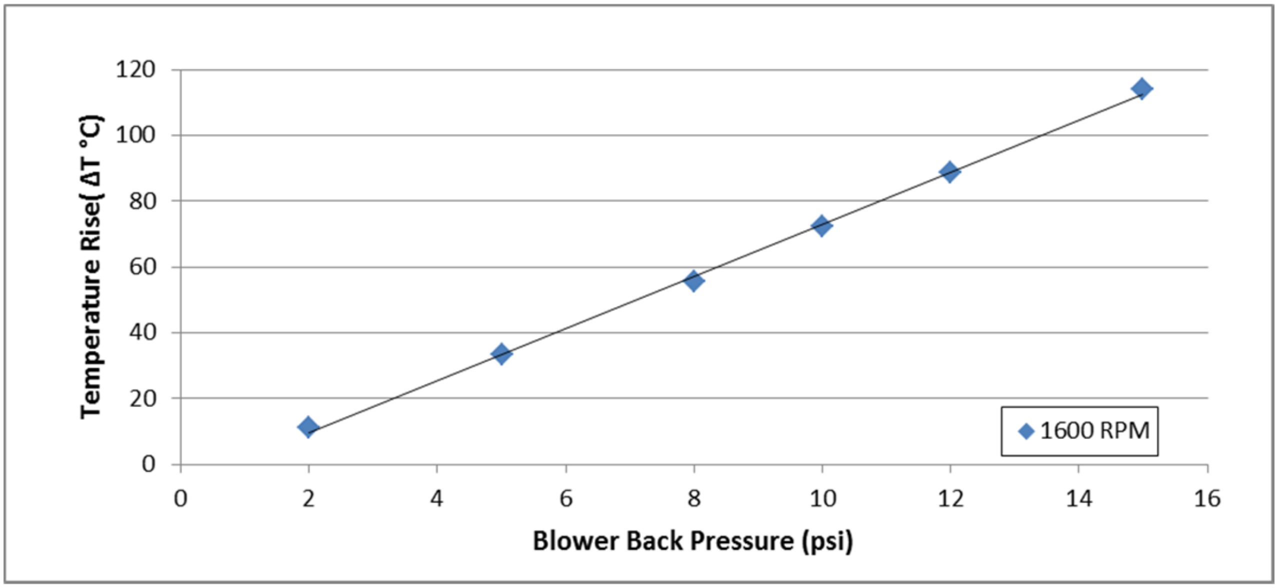


Figure 35. Flue Gas Temperature Increase as a Function of Adsorber Blower Back Pressure.

Immediately following the blower is a shell and tube flue gas cooler, E-101, which utilizes approximately 80 gpm of water to cool the flue gas to the operating temperature of the adsorber, 40°C. This cooler has the ability to cool inlet flue gas from the minimum design temperature of 30°C up to the unaltered outlet temperature of the blower. The cooler and all downstream equipment and piping are constructed of steel.

Downstream of the flue gas cooler is pilot unit bypass ductwork, which has the capacity to direct all flue gas around the pilot unit and return it to the stack. This bypass is utilized during pilot start-up, shut-down, and during operation to regulate the flow through the adsorber.

4.2.3 Adsorber and Adsorber Particulate Control

Flue gas enters the base of a three-stage fluidized bed after exiting the flue gas cooler. The adsorber is a vertical, cylindrical vessel constructed of carbon steel, as illustrated in Figure 36. Flue gas enters the bottom of the lower bed and proceeds upwards at sufficient velocity to fluidize the sorbent. Sorbent BN has a minimum fluidization velocity of 0.86 ft/sec, as discussed in Section 3.6.8. Above each bed, there is a disengagement zone where gravity carries most of the sorbent back into the bed immediately below the zone. Flue gas then flows through a distribution plate and into the next fluidization stage until the low-CO_2 gas exits the top of the adsorber.

Flue gas temperature at the inlet of the adsorber is controlled to 40° C (104° F) for sorbent BN. The temperature is affected by a number of factors, including the flue gas blower inlet temperature, the flue gas blower back pressure, and the flue gas cooler water flow rate. These parameters can be adjusted to set the inlet flue gas temperature at the adsorber. The pilot blower inlet temperature can be manipulated via the slurry cooler built into the SO_2 polishing scrubber and the blower outlet temperature is due to the work performed on the flue gas when it is compressed to overcome the pressure drop associated with the adsorber, as discussed in Section 4.3.2.

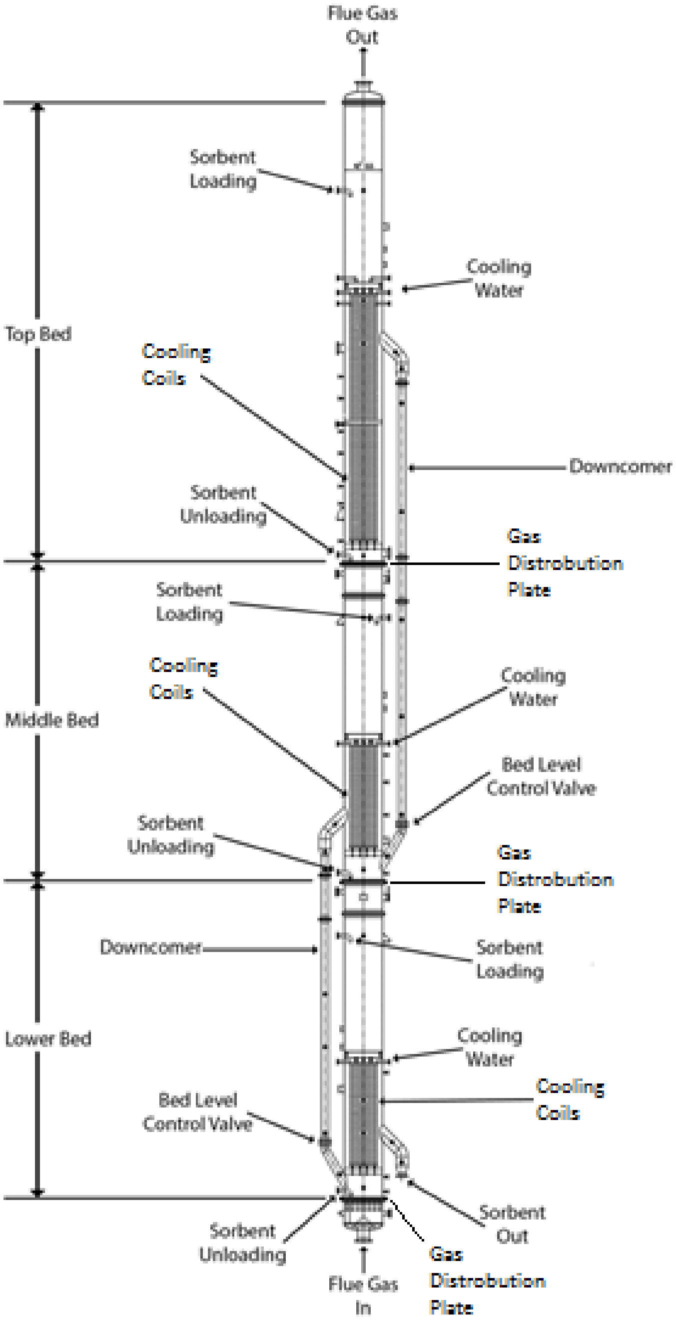


Figure 36. Three Stage Adsorber Design.

The adsorber was designed with water-cooled coils in each bed to maintain isothermal operation of 40°C (104°F). The surface areas of the cooling coils within the bottom, middle, and top beds are 430 cubic feet (cf), 430 cf, and 1014 cf respectively. The amount of cooling needed is dependent upon the flue gas cooler outlet temperature, the operating temperature of the regenerator, and the sorbent heat of reaction.

Densely packed CO_2 loaded sorbent exits the bottom of the pilot adsorber through a dip leg and slide gate valve. This adsorber outlet valve determines the rate at which sorbent circulates within the system as the regenerator empties via an overflow. The sorbent mass flow rate range ranges from 0 to 33,000 lb/hr. The sorbent within the dip leg must be kept in dense phase to minimize flue gas introduction into the regenerator. Approximately 78 SCFM of CO_2 from the CO_2 product stream is used to pneumatically convey the loaded sorbent to the top of the regenerator.

CO_2-deficient flue gas exits the adsorber through a two stage cyclone, CYC-101. As the flue gas tangentially enters the cyclones, the heavier, denser sorbent particles are carried to the walls and directed to dip-legs at the bottom of the cyclones, while the flue gas leaves from the top of the device. This cyclone is designed to remove particulate from the gas stream at 99.99% efficiency within the designated sorbent particle size distribution, as discussed in Section 3.6.8 and show in Table 13. The cyclone is rated to handle up to 2464 ACFM of flue gas with a dust loading up to 21,377 lb/hr at this efficiency. The solids that have been removed from the gas stream via the cyclones are returned to the adsorber through the dip-legs.

The flue gas that leaves the cyclones is routed to a pulse-jet fabric filter which removes very fine sorbent that may have been damaged through attrition as well as any sorbent which may be carried through the cyclones in the event of an upset in operation. The fabric filter is capable of handling 3,600 ACFM of flue gas with an air/cloth ratio of 5.6:1. The fabric filter is equipped with a flame arresting deflagration panel to act as a vent in the event of a deflagration. In the pilot unit, the low CO_2, particle-free gas is then returned to the host unit.

4.2.4 Cooling Water System

The pilot was designed with a once-through cooling water system using cooling water extracted from the river that flows past the host site. The river temperatures from June 2010 through July 2011 are shown in Figure 37. Although the cooling water system could be modified with a chiller to provide additional range for cooling water inlet temperatures, such a cooler was not available for this project. Thus, the operating period of the pilot was constrained to spring and fall months to ensure sufficient cooling capacity for pilot operations while using sorbent BN. Furthermore, the pilot was not constructed for operation during freezing weather and, therefore, operation during winter would not be practical.

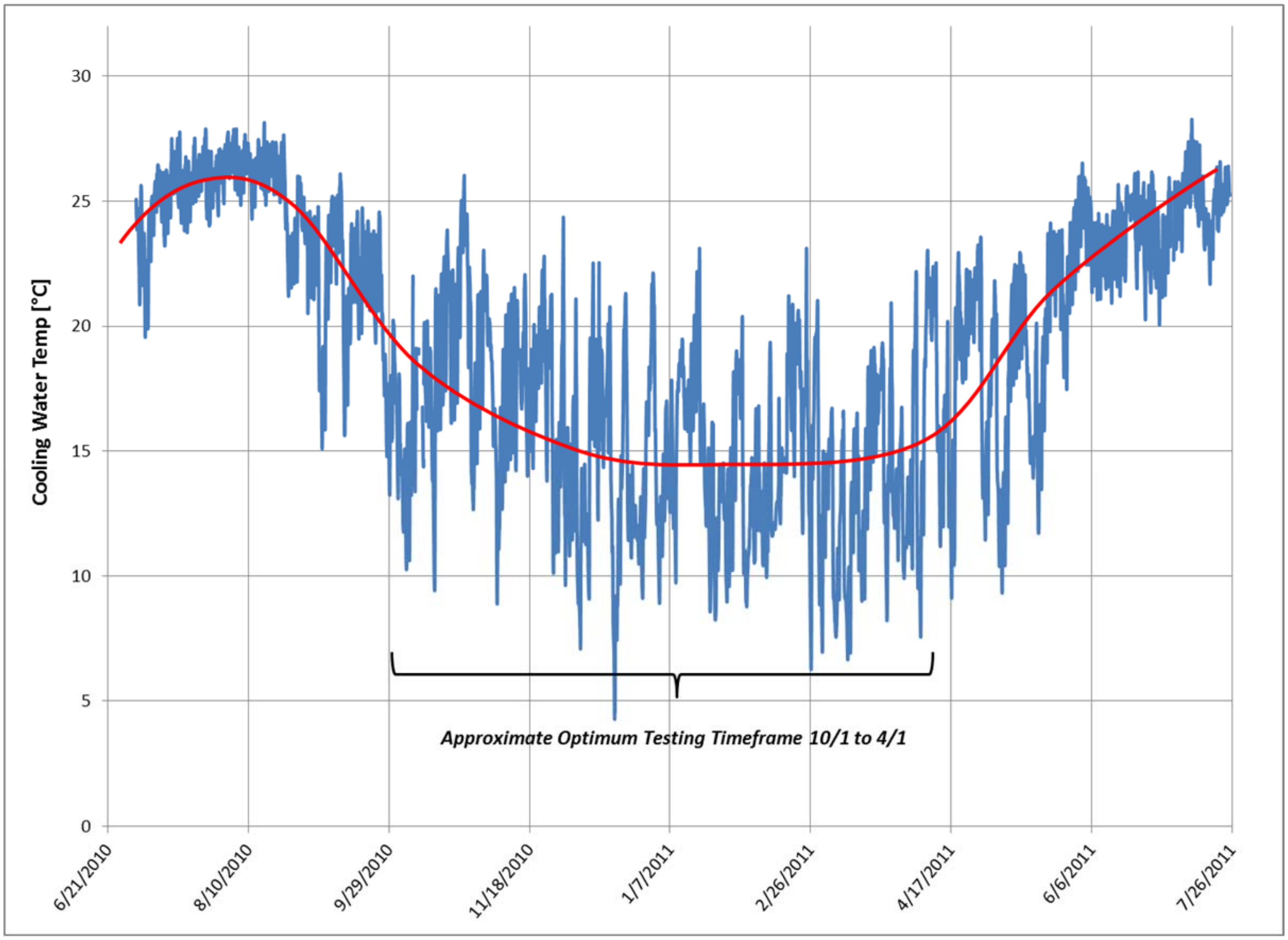


Figure 37. Temperature of River Water Used for Pilot Plant Cooling Water.

The river water used for the pilot is taken from and returned to the generating units WFGD make up water header. This water is untreated filtered and untreated. The CO_2 pilot utilizes a “witch’s hat”-style strainer to remove large debris prior to the entrance of the cooling water pump, P-102. The cooling water pump is designed to operate at 1285 gpm, 80 psig, with a head of 191 ft. The pump is rated to provide 1520 gpm, 88 psig, with 244 ft of head. This pump provides all cooling water for the flue gas and CO_2 coolers as well as the scrubber slurry cooler and the adsorber coils.

4.2.5 Regenerator and Regenerator Particulate Control

The regenerator consists of a single stage fluidized bed, illustrated in

Figure 38, which utilizes a condensing heat exchanger to condense steam to heat the sorbent and gas within to the required regeneration temperature of 120°C. Indirect heating allows for better control of the system to prevent hot spots and potential damage to the sorbent as a result of exposure to high temperatures. The steam supplied to the regenerator’s heat exchanger has a maximum capacity of

3,770 lb/hr at the design conditions of 149°C (300°F) and 86 psig. The heat exchanger coils have a surface area of 1,188 ft^2. The steam supplied to the regenerator in the pilot has the ability to be supplied to the heating coils at a minimum of 3 psi up to the supply pressure of approximately 500 psi. The steam temperature supplied to the coils can range from the respective condensation temperature at a given pressure up to the supply temperature of approximately 371°C.

As loaded sorbent is heated in the regenerator, CO_2 is released from the sorbent, which then passes through a two-stage cyclone and a small fabric filter as was described in Section 4.3.3. Approximately 400 SCFM of the CO_2 is recycled to use as regenerator fluidization gas. In a commercial application the balance of CO_2 would be transferred to a compression plant, while the pilot returned it to the plant stack.

In addition to the blower for the adsorber, another blower, blw-103, is necessary to provide the motive force for the fluidizing CO_2 in the regenerator. The temperature increase in the recirculated gas entering the regenerator as a result of the compression is shown in Figure 39. Blower 103 was designed to operate with an outlet temperature and pressure of 120°C and 5.9 psig. This blower is operationally limited by the outlet temperature and is rated for a maximum continuous operational outlet temperature of 190°C (375°F). Prior to entering this blower the fluidizing CO_2 is passed through a shell and tube heat exchanger, E-102, similar to the flue gas cooler. This cooler is designed to utilize 60 gpm of cooling water to cool the CO_2 to an exit temperature of 40°C. Cooling the CO_2 removes excess moisture and conditions the gas before it is compressed and subsequently heated in the blower.

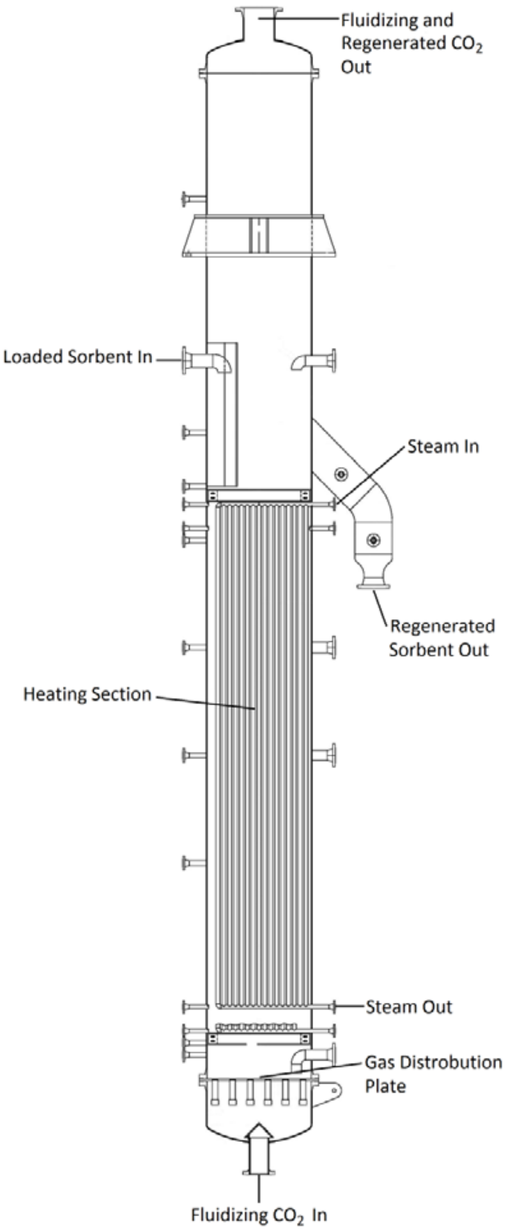


Figure 38. Regenerator Design.

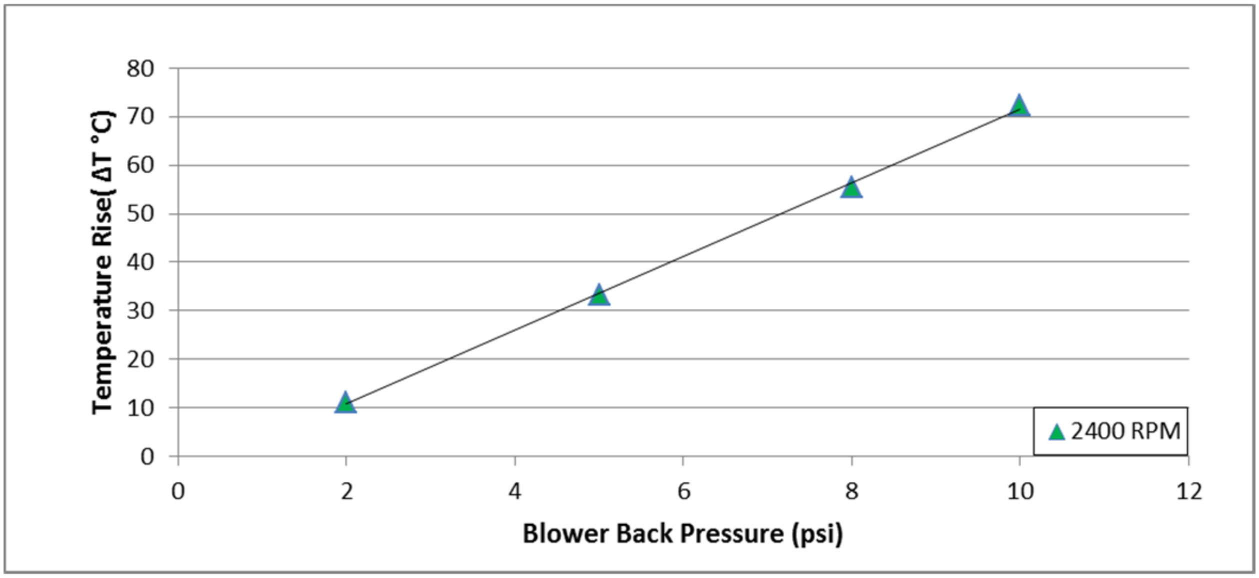


Figure 39. Gas Temperature Increase as a Function of Blower Back Pressure in the Regenerator.

Utilizing both the temperature increase as a result of the regenerator blower back pressure and the steam in the heat exchanger, sufficient heat duty is available for many practical temperature-swing process examinations.

As mentioned in Section 4.3.3, the regenerator level is designed to be held constant by utilizing an overflow design for emptying the regenerator. Regenerated sorbent passing through the overflow is delivered to a cyclone, CYC-104, at the top of the pilot unit via pneumatic conveyance using approximately 292 SCFM of 120°C CO_2. The cyclone separates the heated regenerated sorbent and delivers it to the top bed of the adsorber through a densely packed dip leg to minimize the amount of CO_2 introduced into the scrubbed gas stream. The conveying CO_2 is returned to the product stream.

4.2.6 Future Use

The 1 MWe pilot facility was designed using characteristics of a single sorbent though it can provide some operational flexibility to accommodate future testing campaigns with alternate sorbents that have different characteristics. The range of operation of major equipment and services has been discussed in this Section, 4.3. In order to accommodate different sorbents it may be necessary to add equipment for cooling water refrigeration depending on the specific sorbent's heat transfer coefficient. It is possible to operate with sorbent s of varying densities due to the operational range of the fluidization gas blowers. The extent to which the pilot may provide flexible operation is fairly broad in terms of flue gas flow rates, but will not be able to exceed 1 MWe equivalent. Turndown has proven to be effective to as low as 200 kWe for stable operation.

Additional information associated with operational issues encountered during testing and recommendations for additional testing with sorbent BN are included in Section 6.6.

4.2.7 Additional Design Factors

The final pilot design incorporated site-specific details including plant interface requirements and permitting requirements. These activities are summarized below.

Plant interface requirements were established based upon consensus between the host plant facility and ADA. Due to some preferences for steam line integration and other plant concerns, small modifications were made to the pilot system to ensure the least disruptive integration of the pilot facility with the host site.

Electrical, steam, water, and other utilities connections were determined and then agreed upon with the host site. The pilot construction company, McAbee Construction, and ADA worked with the plant to carry out the necessary utility connections and plant integration.

While designs were being completed and plant interface requirements were worked out, ADA solicited bids for subcontracting services to provide the necessary equipment, materials, fabrication, and installation work required for pilot plant implementation. Potential vendors and suppliers were engaged and specific scopes of work were created for selected vendors.

Permits, environmental and construction, and NEPA clearance as required by DOE and the host site were obtained. A Go/No Go decision was made in conjunction with a continuation application for funding before construction commenced. As part of the application and necessary permitting requirements, multiple documents and permits were negotiated such as the following.

- Host site agreement.
- Environmental questionnaires for the host site and other project team sites or activities not addressed in prior submittals.
- Environmental and construction permits.
- Topical report on sorbent analysis and selection.
- Preliminary test plan.
- Complete detailed design package with quotes and delivery dates for selected service contractors and major equipment purchases, plus all other expenditures documented to support budget requests for Budget Period 2.
- Updated Project Management Plan with schedule and Statement of Project Objectives.
- Phase 1 costs to date and estimated costs to complete.

5. Procurement and Construction Activities

5.1 Procure & Manufacture Sorbents

ADA contracted with a confidential third party to produce two 10,000 lb batches of raw sorbent. The quantity of sorbent ensured that if one batch of pilot sorbent was contaminated or destroyed during pilot operation, that a separate batch of sorbent would be available to facilitate continued testing.

Some post-manufacturing processing was required to prepare the raw sorbent for testing. The relatively small amount of sorbent required by ADA for pilot testing made it cost prohibitive to manufacture the sorbent at the size distribution needed for pilot operation. This sorbent has the ability to be produced at specified particle sizes and post-manufacturing processing would not be required of a commercial product. A large batch trial was conducted to dry the sorbent to less than 12% moisture in a vacuum dryer and the results were not acceptable. A second large batch trial attempting to dry the sorbent in a fluidized bed dryer was successful. After achieving acceptable drying results, the sorbent designated for pilot testing was processed to the specified particle size distribution via pin milling and screening.

ADA worked conducted quality assurance tests to ensure the sorbent had been manufactured to the correct particle size distribution discussed in Section 3.6.8. Samples of the sorbent were sent to ADA and tested in ADA's laboratory during every step of the processing to ensure the sorbent adsorption properties were not altered. Quality control sampling took place to ensure similar performance characteristics between batches. Specifically, several different lots of the sorbent were analyzed in ADA's laboratory to ensure the 1 MWe pilot sorbent was suitable and exhibited similar characteristics of previously analyzed batches.

5.2 Procure and Construct Pilot-Scale Equipment

ADA selected a McAbee Construction Inc. due to their tremendous amount of experience in module fabrication to execute the following work packages:

- Fabrication of Adsorber and Regenerator Vessels
- Fabrication of Pilot Modules
- On-Site Installation Activities
  - Foundation, UG Utilities
  - Piping Installation
  - Module Installation and field erected structural steel
  - Electrical Scope of Work
  - Piping and Equipment Insulation

In an effort to reduce costs associated with the fabrication and installation of the pilot unit, ADA conducted several conference calls between project participants. The construction company believed costs could be further reduced by limiting site construction activities and increasing the module scope of work by pre-assembling all equipment to the greatest extent possible at the module shop. Process design and the equipment layout were evaluated to ensure on-site activities were minimized.

ADA hosted an engineering, procurement, and construction (EPC) kick-off meeting at the ADA corporate headquarters in Highlands Ranch, CO. This meeting served to set the expectations and lay the project management ground work for the EPC tasks. In addition ADA participated in a host site discovery meeting. This meeting was hosted by ADA and the host site, and had participants from many project stake holders. During this meeting measurements and drawings were procured for pilot utility routing design.

Project team members from ADA, the host site, and Technip conducted a hazard and operability study (HAZOP). During the HAZOP, the project team evaluated several operational deviations (e.g., no flow, less flow, more flow) for each process line. During this exercise, the project team discussed the consequences of such actions and identified the necessary safeguards which needed to be installed to ensure safe operation of all ancillary systems included in the pilot unit. At the conclusion of this exercise, the project team identified several areas which needed to be addressed (e.g., ambient monitoring of CO_2 and oxygen near sorbent unloading areas). The HAZOP was beneficial to the project and useful for evaluating new technologies.

5.2.1 Procure Pilot Scale Equipment

Before fabrication began, ADA worked closely with Technip to finalize vendor procurement documentation and equipment quotes. Revised quotes and equipment data sheets were used for the procurement specifications and were part of the procurement contractual documentation with purchase order issuing.

Procurement for instrumentation, all large equipment, and critical path components began at the time McAbee was provided with the structural information needed to develop a materials take off (MTO) for the pilot structural steel order, to ensure delays were minimized. All equipment was shipped to McAbee's facility in Tuscaloosa, AL. Equipment procured for the project was inspected and inventoried as it was received.

5.2.2 Finalize Fabrication and Construction Work Packages

An important component of detailed design of the 1 MWe pilot was the 3D model. The 3-D model was used to support fabrication. Therefore, the 3-D model was reviewed at several different stages of completion. An example of an overall 3-D rendering of the pilot is shown in Figure 40. The adsorber is the tallest module in the figure. A 30% complete model review for the 1 MWe pilot was conducted between ADA, Technip, and McAbee. From this meeting the project team was able to decide on major equipment locations, which facilitated pipe routing for the 90% complete model review. Once the model was completed the fabrication and construction work packages were able to be put together.

Once the fabrication and construction work packages were finalized a construction schedule and contract coordination meeting was held at McAbee's office in Alabama. The purpose of this meeting was to coordinate schedule and contractual details while the project transitioned from design engineering to construction efforts.

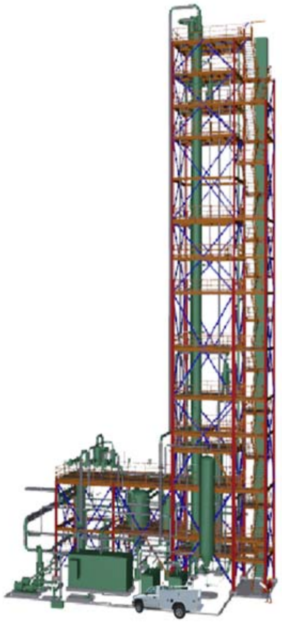


Figure 40. Sample 3-D Rendering of the Pilot.

In January 2013, a construction kick-off meeting was held at the host site. The purpose of this meeting was to coordinate schedules and communication between all project stakeholders, plan construction activities ensuring all concerns are addressed, and evaluate any impacts.

5.2.3 Fabrication and Construction of Pilot Scale Equipment

Fabrication of the pilot was performed in Alabama. The vessels and structure were constructed in a fabrication shop and module yard. When the modules were completed, they were loaded on a barge and transported upriver to the host site. A photo of the modules during construction is shown in Figure 41. A photo of the pilot during transport on the barge is shown in Figure 42.

After arrival at the host site, the modules were unloaded and erected on a concrete pad at the host site. After erection of the modules, wiring and plant tie-ins were completed. Activities at the host site were supervised by ADA personnel.

Figure 41. Photo of Pilot Modules under Construction.

Figure 42. Pilot during Barge Transport.

5.2.4 Pilot Installation

While fabrication of the pilot vessels and structures was being performed, installation of foundations and drainage systems, structural support steel and inlet/outlet ductwork with isolation valves, and required utilities (electrical, compressed air, steam) was completed at the host site. ADA worked to provide the necessary compressed air and subsystems for installation at the pilot while contractors performed the foundation work, electrical work, and tie-in to the plant steam system and duct work.

Once the pilot was delivered to the host site, it was installed on the prepared foundations. A photo of the pilot during erection activities is shown in Figure 43.

Figure 43. Pilot Erection at Host Site.

Installation progressed smoothly throughout and substantial completion of pilot mechanical and electrical activities was completed within a year of the start of module fabrication. ADA's onsite construction manager worked with the construction company's installation team during all construction and installation activities.

Above-ground piping and electrical installation activities commenced once the modules were installed at the host site. ADA's onsite construction manager worked with the construction company's installation team as well as plant personnel to finalize installation activities during first quarter of 2014. Electrical scope of work completed included conduit and cable tray installation, plant interface wiring, interconnecting wiring between field devices and control building, and verification of point to point wiring checks.

6. Pilot-Scale Operation and Evaluation

6.1 Pilot Testing Plan

A series of parametric and a continuous operations period was planned to validate the following key performance parameters. A review of these parameters is provided in Section 6.4.4.

- Reactor pressure drop
- Successful isothermal operation of the adsorber and regenerator
- CO_2 delta loading
- H_2O delta loading
- Combined enthalpy of reaction/adsorption of CO_2 and H_2O
- Reaction kinetics of sorbent BN
- CO_2 partial pressure in the regenerator
- Optimize process variables:
    - Temperature (adsorber and/or regenerator)
    - Sorbent Circulation Rate
- Identify sorbent bed height required to maintain 90% CO_2 capture
- Confirm key fluid bed design parameters:
    - Bed density
    - Heat transfer coefficient (cooling in adsorber, heating in regenerator)
    - Riser lift gas requirement
    - Standpipe aeration gas requirement
- Assess the experimentally predicted number of adsorber stages (3) required for 90% CO_2 removal with BN sorbent
- Confirm the overall feasibility of using solid sorbent BN for CO_2 capture from power plant flue gas in a staged fluidized bed adsorber and fluidized bed regenerator

6.1.1 Parametric Testing Plan

A series of parametric tests was planned to demonstrate 90% CO_2 capture at the design conditions and determine sensitivity to flow and temperature. Approximately two weeks of parametric tests were scheduled to be completed once the system was operating stably on flue gas and circulating sorbent without the introduction of steam.

With flue gas circulating through the system, sorbent was introduced and the system was filled. To avoid any condensation, cooling water would be pumped through the cooling coils only after the exothermic CO_2 loading of the sorbent had increased the temperature of the system to at least 40°C. The heat exchangers responsible for removing heat in the adsorber would be assessed; specifically, the ability of the system to operate isothermally under the design operating conditions.

Once stable operation and sorbent circulation were demonstrated and the sorbent had been loaded with CO_2, the regenerator temperature would be incrementally increased via the addition of steam for indirect heating. The temperature of the regenerator would be slowly increased to avoid exposure of the sorbent to high O_2 concentration at high temperature, which could lead to oxidation of the amines. The steam usage would be continuously measured and, before increasing the regeneration temperature, the successful performance of the heat exchanger in the fluidized bed regenerator would be confirmed.

Once the regenerator temperature were above approximately 70°C the CO_2 laden sorbent would begin to regenerate and, therefore release CO_2. However, the working CO_2 capacity was expected to be unacceptably low at this regeneration temperature. The system would be allowed to achieve steady state operation and after the CO_2 mass balance had been closed, the regeneration temperature would be increased. The system would then be allowed to reach steady state conditions again at several different regeneration temperatures. The regenerator temperature was to be increased to 120°C unless 90% CO_2 capture were achieved at a lower regeneration temperature.

There are several key operating parameters that can be varied to ensure that 90% CO_2 capture is attained, including the following.

- Adsorption temperature.
- Regeneration temperature.
- Sorbent circulation rate.
- Bed height in each of the staged fluidized beds.
- Flue gas volume into the adsorber.

The goal of the parametric tests was to identify the optimal operating conditions prior to initiating the continuous performance-testing period.

6.1.2 Continuous Performance Testing Plan

The continuous testing subtask was designed to obtain sufficient operational data on removal efficiency to determine if 1) there is a degradation in the CO_2 removal performance of the sorbent and 2) to monitor process parameters such as temperatures, pressures, and sorbent attrition rates at an operating condition that is determined from results of the parametric testing and process modeling sensitivity analyses as "optimum" by ADA and subcontracted engineers and approved by DOE/NETL, and the rest of the project team. The 1 MWe pilot was operated starting at the optimal conditions identified during the parametric testing. Once the condition of 90% CO_2 capture had occurred, the plan was to maintain consistent operating conditions throughout the duration of the test, unless the onsite team decides that another set of conditions would be superior based on heat duty, pressure drop, etc. The quality of the concentrated CO_2 stream was monitored during continuous testing.

6.1.3 Test Methods

Gas Analysis

The 1 MWe pilot was instrumented to allow for detailed analysis of temperatures, pressures, gas flow rates, CO_2 concentrations, and moisture levels throughout the system. A series of gas analyzers was utilized to measure the gas at the adsorber inlet, outlet, and the regenerator outlet; these sampling points allowed for a mass balance to be calculated.

Sorbent Analysis

Samples of the sorbent were collected throughout the duration of 1 MWe pilot testing. The sorbent from different locations was analyzed periodically and tested in the laboratory to assess if any degradation had occurred and for analysis to monitor any uptake of TCLP Heavy Metals. The sorbent was collected from the adsorber and the regenerator. Results are presented in Sections 6.3 and 6.4.

Gas Sampling Locations

The gas sampling locations are shown in Figure 44 and described in Table 17.

The host site also provided the flue gas composition, specifically the SO_2 concentration, in the flue gas upstream of the 1 MWe polishing scrubber. The intermittent analyzers for the CO_2, O_2, and H_2O were shared, and were thus operate intermittently at any one sampling point. CO_2 and O_2 analyzers for locations 3A and 6A were remotely located in a shelter; all other analyzers were located at the sample points. There was an additional CO_2/O_2 analyzer other than those listed in Table 17. This additional analyzer was placed outdoors near the bottom of the baghouses to measure ambient CO_2 and O_2 concentrations for safety purposes.

Table 17. Gas Sampling Locations.

Sampling Point	Gas(es) Sampled	Frequency of Sampling	Analyzer Location	Used for Mass Balance
3A	CO_2,O_2, SO_2, CO, NO, NO_2	Continuous	Adsorber Inlet	Yes
4A, 5A	CO_2,O_2, SO_2, CO, NO, NO_2	Intermittent	Fluidized Bed Overhead	No
4B, 5B	H_2O	Intermittent	Fluidized Bed Overhead	Yes
6A	CO_2,O_2, SO_2, CO, NO, NO_2	Continuous	Adsorber Outlet	Yes
6B	H_2O	Intermittent	Adsorber Outlet	Yes
8A	CO_2,O_2, SO_2, CO, NO, NO_2	Continuous	Regenerator Outlet	Yes
8B	H_2O, O_2	Intermittent	Regenerator Outlet	Yes

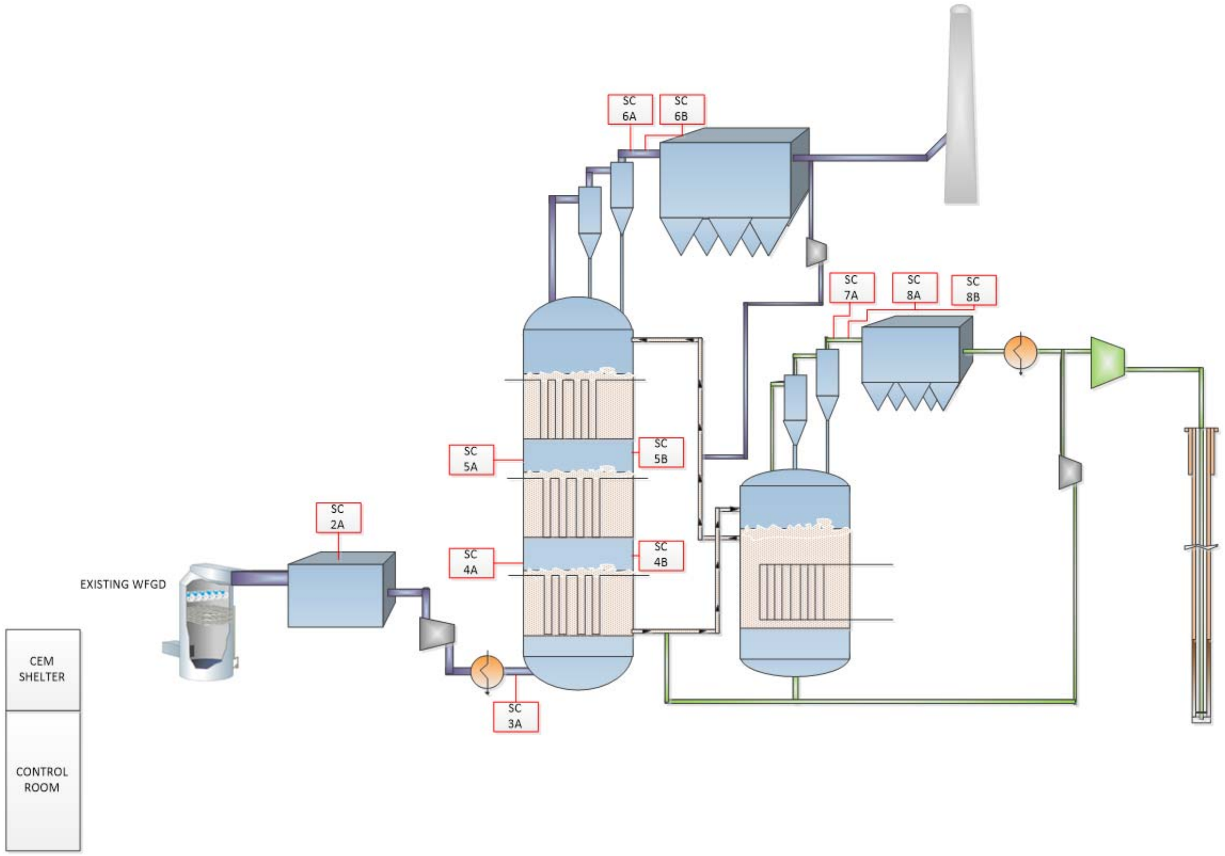


Figure 44. 1 MWe Pilot Gas Sampling Locations.

6.2 Commissioning and Start-up

ADA prepared a startup and commissioning plan detailing the sequence of commissioning activities to ensure all equipment operates in a safe and reliable manner. As each major equipment item was built, a factory acceptance test (FAT) was conducted at the equipment manufacturer to ensure the equipment would operate at design specifications.

The commissioning plan detailed the startup tests to monitor and measure the performance boundaries of the system with all the equipment in place. The primary areas of focus were scrubber functionality, flue gas flow, pressure drop, cooling/heating capability and control within all vessels, HMI control, control loop tuning, and data collection capability. The commissioning period was to last for a period of two months, an aggressive schedule that allowed for limited opportunity to correct any significant equipment operating issues.

ADA began operation of the ADAsorb pilot on September 29, 2014. Upon startup, sorbent was loaded into the system and sorbent circulation was demonstrated and the sorbent control loops were tuned. To avoid any condensation, the cooling water was pumped through the cooling coils only after the exothermic CO_2 loading of the sorbent had increased the temperature of the system to at least 40°C (104°F). The heat exchangers responsible for removing heat in the adsorber were assessed; specifically, the ability of the system to operate isothermally under the design operating conditions. Once these conditions were met, the adsorber inlet flue gas flow rate was gradually adjusted to a rate representative of a 1 MWe.

Upon achieving an adsorber inlet flue gas flow rate representative of 1 MWe, steam was introduced into the regenerator. The regenerator temperature was slowly increased via the addition of steam for indirect heating. The temperature of the regenerator was slowly increased to avoid exposure of the sorbent to high O_2 concentration at high temperature, which could lead to oxidation of the amines.

6.3 Pilot Testing Periods

Pilot testing was defined as three distinct periods: Round 1 Parametric Testing from October 10 and November 11, 2014, Round 2 Parametric Testing in spring of 2015 between April 4 – May 4 and May 15 – June 25, 2015, and Continuous Operation at 90% CO_2 Capture for 20 hours starting on June 25, 2015. Operation during these periods is discussed in this section. An analysis of results from testing is presented in Section 6.4.

6.3.1 Round 1: Parametric Testing

Round 1 of parametric testing was conducted between October 10 and November 11, 2014. Several operational issues were encountered that required system repairs, including damage to deflagration panels in the baghouse resulting unstable operating conditions and a large surge of sorbent carrying over into the baghouse, instrument grounding issues that caused unreliable signals and difficulty in operating the system, demister pad pluggage in the polishing scrubber and associated increased pressure drop and caustic carryover to the blower. These issues and other project delays associated with a plant outage are detailed in the section below.

Round 1 of parametric testing began with the introduction of steam at 16:30 on 10/10/2014. Between the time of steam introduction and 4:10 on 10/11/2014, 11 hrs and 40 min, the sorbent temperature within the regenerator was increased from 40.5°C to the first parametric set point of 80°C. Once the sorbent became saturated with CO_2 and before the regenerator was operating at temperature and functioning to reduce the CO_2 concentration on the sorbent, the inlet CO_2 was approximately equal to the outlet CO_2.

An example of data collected during regenerator heat-up is provided in Figure 45. Unfortunately, before stable operating conditions in the system could be achieved, an upset condition occurred that resulted in a system shut-down. Specifically, immediately upon achieving the 80°C regenerator temperature, the adsorber fluidized bed became unstable, causing a surge of sorbent to exit the adsorber. The sorbent back-flowed through the sorbent return dip leg of the cyclone located at the exit of the adsorber that was designed to separate sorbent from CO_2-lean flue gas, return the sorbent to the adsorber and pass the flue gas to the baghouse. When sorbent flowed backward through the return dip leg, the cyclone was ineffective and sorbent was carried directly into the baghouse. Deflagration panels were located in the baghouse that were designed to fail open in the event of an explosion. The excessive sorbent entering the baghouse exceeded the 1.0 psig burst pressure of the deflagration panels. The pilot was forced to shut down until the panels could be repaired.

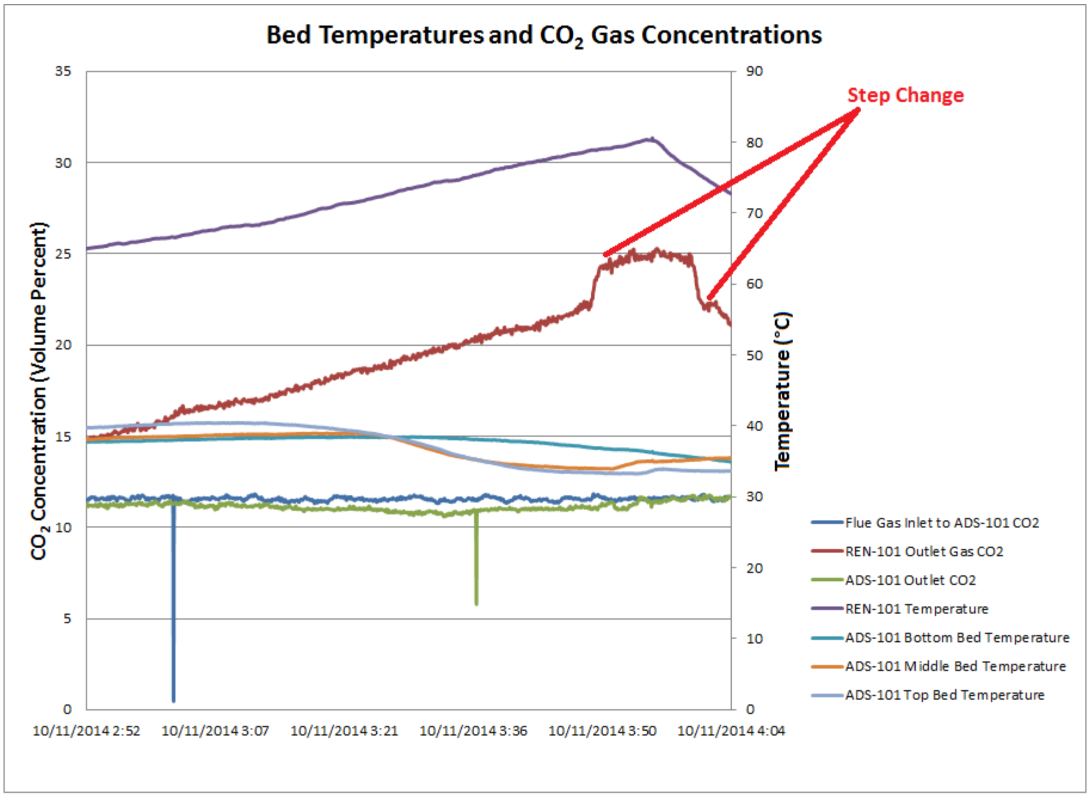


Figure 45. Round 1 Parametric Testing Data Example.

The deflagration panel was replaced and on 10/25/2014 the pilot was at steady-state flow and sorbent circulation, ready to reintroduce heat to the regenerator and resume parametric testing. Unfortunately, the host site went into an unscheduled outage at the same time so the pilot had to be systematically shut down. On 10/29/2014 the host site began start up and ADA was given permission to resume operation of the pilot unit. Upon startup of the pilot unit's flue gas blower, BLW-101, an internal coupling sheared and the pilot was once again required to shut down.

With the blower coupling repaired at 14:15 on 11/3/2014 the pilot was once again ready to reintroduce steam and resume parametric testing. The first set point of 80°C was achieved that day at 17:49 and steady operation of the pilot at this condition began. During the operation of the pilot at this condition, it was noted by operators that large fluctuations in process values was occurring which made steady operation very difficult. Over the course of the following two days parametric testing at one other temperature was attempted with minimal success due to the apparent inability to understand the interrelations between process variables which made steady state operation practically impossible.

At midnight on 11/6/2014 the steam system was once again shut down. The desuperheater, which both cools and reduces the pressure of the steam for the pilot plant, was subsequently disassembled and inspected due to per the recommendation of the manufacturer due to the inability to accurately control steam temperature and pressure within the regenerator. No issues were discovered with the desuperheater. It is possible that the steam temperature measurements were erroneous, as described in more detail below.

During this period of operation it should also be noted that the pressure at the outlet to the scrubber/ inlet to the flue gas blower had been steadily decreasing and no recommendations provided by the scrubber manufacturer to reduce this pressure through demister washing had been effective.

Between 11/6/2014 and 11/10/2014 it was discovered that local instrument displays were displaying values different than values read within the programmable logic controller (PLC) panel via multi-meter and on the operator human-machine interface (HMI). Prior to startup all instrumentation was calibrated and all values were verified to be consistent at all three locations. Through troubleshooting it was uncovered that faulty grounds existed within the PLC cabinet. During operation of the pilot, it is believed that static generated due to fluidizing the sorbent material enhanced the issues. The instrument values operators were using for pilot control were unreliable and often erroneous. The discovery of the grounding fault issues were likely the cause of many of the control issues experienced and noted by the operators and seemingly erratic behavior of the system, such as the step change illustrated in Figure 45, and the desuperheater temperatures.

Steam was reintroduced to the system on 11/10/2014 after ground-fault issues were addressed. Before the system could get to the target set point, it was noted that the pressure drop across the flue gas blower was approaching the maximum differential pressure of 15 psig corresponding to the high vacuum at the outlet to the scrubber/ inlet to the flue gas blower, such as that illustrated in Figure 46. In addition to the high vacuum at the inlet the blower was also beginning to make unexpected noise.

Figure 46. Flue gas blower (Blower-101) Inlet Pressure from 10/3/2014 to 11/11/2014.

Upon inspection, the scrubber demister pad was found to be completely fouled and, due to the high vacuum, the supports had broken as shown in the photo in Figure 47. The fouled demister resulted in carryover of caustic material into the flue gas blower, ultimately resulting in the blower seizing during shutdown which required extensive rebuild and repair. During inspection of these components the ambient temperature remained below freezing for several days resulting in damage to cooling water system flanges due to ice formation.

Figure 47. Fouled Demister Pad.

The pilot was online flowing sorbent and flue gas for a total of 21 days during start-up and this first round of parametric testing before it was shut down as a result of the blower damage, demister failure, and freezing ambient temperatures. The system was secured for the winter months and necessary repairs beginning on 11/11/2014.

Pilot testing and collection of data representative of equilibrium operation was limited as a result of controls issues, problems with carryover from the SO_2 scrubber, host unit shut down, and freezing ambient temperatures. However, during the 2014 testing period, operators were able to establish consistent sorbent cycling through the system. Unfortunately, instrument and panel grounding and wiring issues affected the integrity of the collected operational data. Therefore, specific results regarding CO_2 capture and regeneration were unreliable for the first attempts at testing.

Following initial parametric testing, on 11/10/2014, a sorbent sample was collected for laboratory evaluation to determine if exposure to flue gas and cycling in the pilot impacted performance or indicated any potential future handling concerns. Testing was completed for particle size distribution, TCLP (EPA method 1311) testing, and CO_2 capacity. These results, discussed further in Section 6.5, indicated that the sorbent was not subject to contamination from metals present in the flue gas and would not be considered hazardous waste for disposal.

During the pilot shutdown, necessary repairs to the blower were performed as well as a re-engineering of the demister pad and demister wash system by the demister manufacturer. Additional modifications including modifications to the flue gas and CO_2 cooler, E-101 and 102, water supply valves, the addition of scales to both baghouse hoppers, and the addition of blowbacks to all of the cyclones were completed during the required shutdown. These modifications facilitated operation of the scrubber without damage to the flue gas blower and allowed for safer and easier operation the remainder of the project.

6.3.2 Round 2: Parametric and Continuous Testing

The pilot was re-commissioned beginning 3/26/2015 and parametric testing recommenced on 4/9/2015. The spring 2015 pilot operation was conducted between 4/3 - 5/4/2015 and 5/15 – 6/26/2015. Due to sorbent carryover into the fabric filters during initial parametric tests during Round 1, parametric testing was initiated at a reduced adsorber inlet flue gas flow rate of approximately 850 ACFM, or 0.5 MWe to better manage any variations in sorbent handling characteristics resulting from heating the sorbent.

During the second round of parametric testing, material handling of the heated sorbent continued to be a challenge. For example, pluggage of the sorbent return dip leg below cyclone-101, the cyclone designed to separate sorbent returning to the adsorber from the regenerator from carrier gas, as well as complete pluggage of the cyclone-102, the cyclone located at the outlet of the regenerator designed to separate the CO_2-rich gas from any sorbent carried out of the regenerator, led operators to shut down the system many times due to upsets and to make modifications. These cyclones are shown in Figure 48.

Additional grounding fault issues were uncovered during operation of the pilot. A third-party instrumentation specialist was brought in to assist in troubleshooting signal interference issues experienced by the operators. These issues resulted from poor wire practices by a third-party electrical contractor during fabrication, which included removing shielding from instrument wiring several feet prior to termination in the controls cabinet. Modifications to the wiring were made, including replacing necessary shielding and assuring adequate grounding. These modifications seemed to resolve the issues.

On June 6th additional modifications were made to the dip leg of cyclone-104, the cyclone that separates sorbent returning from the regenerator from carrier gas, within which pluggage and slug flow out were occurring. Previously two vibrators had been mounted to this 8-inch steel pipe in conjunction with increasing fluidizing CO_2 where the dip leg meets the adsorber. The installation of the vibrators did not resolve the flow issues. In place of the vibrators, fluidization ports were installed in the dip leg into which nitrogen (N_2) could be metered, as illustrated in Figure 48. Once the fluidizing nitrogen was installed and tuned to between 1 to 2 ACFM per location material handling issues were minimized.

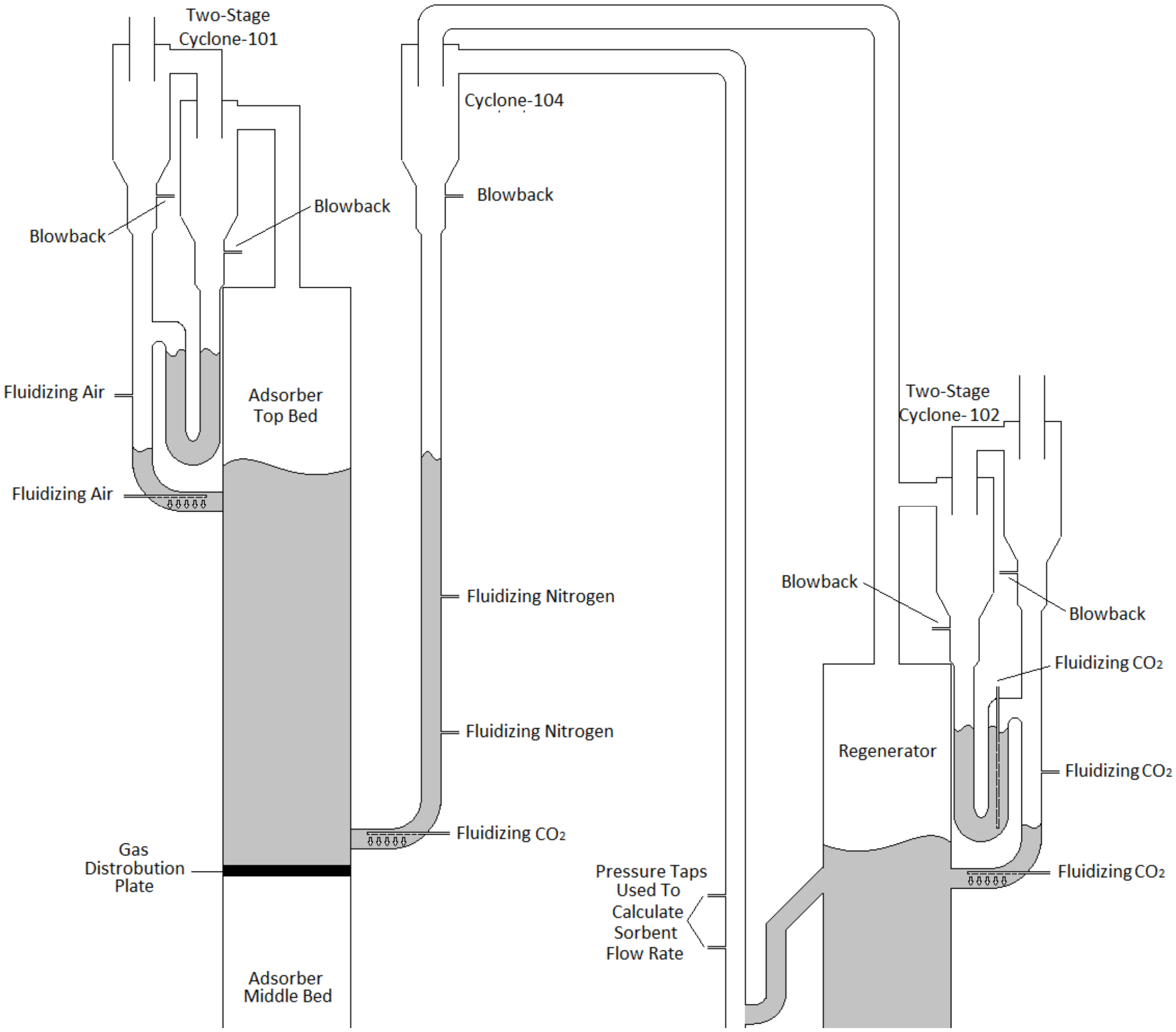


Figure 48. Adsorber and Regenerator Connection Components.

On-site operators noted several indicators that, during some periods of operation, the measured sorbent circulation rate may have been incorrect due to sorbent potentially being recirculated back to the top of the regenerator by passing through the partially clogged cyclone-104 (sorbent return to adsorber from regenerator). Due to the placement of the pressure taps used to measure sorbent circulation rate, as shown in Figure 48, all sorbent exiting the regenerator is assumed to be returned to the adsorber. Using only these pressure taps, any sorbent bypassing the adsorber by recirculating into the regenerator through cyclone-104 will not be accounted for. The theory that some recirculation was periodically occurring was supported by large amounts of sorbent completely clogging the first stage of cyclone-102, the gas exit from the regenerator, leading to sorbent and CO_2 back flowing through the cyclone-102 dip leg and second stage, and carrying large amounts of sorbent into the fabric filter.

Fine tuning of the regenerator outlet CO_2 fluidization valves and the cyclone-104 nitrogen fluidization valves was performed on 6/12/2015. CO_2 capture data before and after fluidization adjustments is shown in Figure 49. This data, combined with the log notes, supports the belief that sorbent flow into the adsorber was periodically limited during operation before the fluidization adjustments. Analysis of the raw data for the two distinct time periods and data sets, pre and post valve adjustment, shown in Table 18, reveals significant changes in some of the results, which supports the lack of sorbent circulation into the adsorber during periods prior to adjusting fluidization.

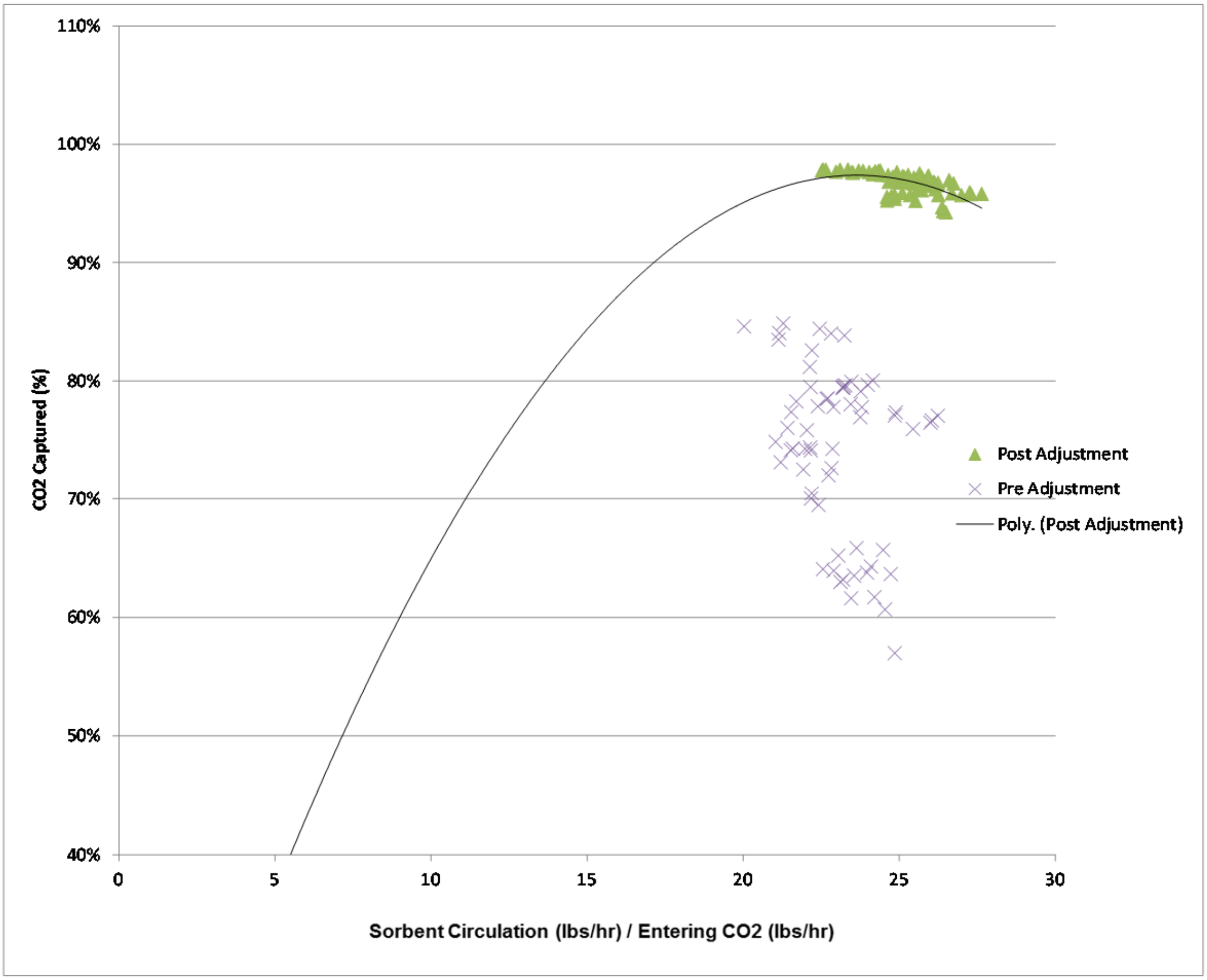


Figure 49. CO_2 Capture Before and After Fluidizing N_2 and CO_2 Valve Adjustments.

Table 18. CO_2 Measurements Before and After Fluidizing N_2 and CO_2 Valve Adjustments.

Tag Description	Units	Pre-Adjustment	Post-Adjustment	%Change
CO_2 entering adsorber	lb/h	474	481	1%
CO_2 exiting adsorber	lb/h	123	15	-720%
CO_2 removal in adsorber	lb/h	351	465	25%
CO_2 removal in adsorber	wt%	74%	97%	24%
Moisture uptake in adsorber	lb/h	45	46	2%
Sorbent working capacity	%	3.2%	3.9%	18%
Total sorbent inventory	lb	9,809	9,885	1%
Sorbent circulation rate entering the adsorber	lb sorbent per lb CO_2	23	25	8%
CO_2 in flue gas entering adsorber	vol%	13.3	13.5	1%
Calculated solid flow in transfer line	lb/h	10,858	12,111	10%
Steam flow rate from plant	lb/h	1,313	1,540	15%
Adsorber exit valve open	%	4.2%	6.9%	39%
Area open in HY-003	in^2	1.4	2.3	39%
Regenerator bed level	ft	18.98	17.34	-9%
Adsorber bottom bed avg. temp.	^oC	37	41	7%
Adsorber middle bed avg. temp.	^oC	39	44	7%
Adsorber top bed avg. temp.	^oC	41	43	3%
Regenerator temp.	^oC	119	121	1%

An increase in the open area of the valve at the bottom of the adsorber, HY-003, increased steam consumption, increased adsorber temperatures, and decreased regenerator bed height post-adjustment are all indicative of more sorbent entering/exiting the adsorber after fluidization adjustments. It is important to note the CO_2 entering the adsorber in lb/hr and volume % remained constant during both time periods as well as the regenerator temperature and total sorbent inventory.

Of interest is the area open within the HY-003 valve itself and how it corresponds to the sorbent circulation rate. Increasing the area open within the valve from 1.4 in^2 to 2.3 in^2, a change of 66%, the circulation rate only registered a 12% change. This is additional support for the belief that sorbent may have been bypassing the adsorber during some of the testing periods.

The top bed at several conditions as well as the middle and bottom bed at two conditions were not able to maintain the desired condition of 40°C. The adsorber bed cooling coils utilized the host site FGD make-up water which was taken from a local river. The pilot unit was designed under the assumption that this water was filtered through a sand filter prior to the pilot extraction location. It was discovered during operation that the water was not filtered and that due to build up on the water pump pre-screen, known as a witch's hat, sufficient water flow to the coils could not be achieved. The fouled pump pre-screen, or "witch's hat", is shown in Figure 50.

Figure 50. Fouled Water Pump Pre-screen.

Although there were indications that some sorbent was bypassing the adsorber, funding limitations and a fixed testing schedule dictated that parametric tests were undertaken to assess the performance of the ADAsorb system. Continuous parametric testing was conducted between 6/6/2015 and 6/24/2015. During this time, the regenerator was held at the design operation conditions of 120°C. The flow rate of the adsorber was varied throughout the period. Some adjustments to the sorbent flow rate were also made. Analysis and discussion of the results are provided in Section 6.4.

6.3.3 Continuous, 90% CO_2 Removal Testing

Although the pilot was operated continuously for most of the month of June 2015, the period designated as "continuous operations" was during 20 hours on June 25, 2015 when the gas flow rate was reduced sufficiently to achieve 90% CO_2 capture. As discussed previously, the combination of CO_2 adsorbed while transporting the regenerated sorbent to the adsorber, and excess sorbent residence time in the adsorber when the sorbent was approaching equilibrium capacity, limited the ability to achieve high capture at reasonable gas flowrates.

6.4 Analysis and Discussion of Pilot Results

6.4.1 First Round of Parametric Testing

Due to the faulty data resulting from grounding issues and the limited data, few conclusions can be drawn from the first round of testing. It was possible to achieve stable sorbent circulation in the three stage ADAsorb system at the design sorbent circulation rate prior to introducing steam to the regenerator and heating to the design temperature of 120°C.

Figure 45 in Section 6.3.1 shows the CO_2 concentration exiting the adsorber trending up as the sorbent temperature within the regenerator trended up, indicating that the regenerator effectively desorbed CO_2. Laboratory testing of sorbent collected from the pilot also indicated exposure to flue gas and cycling did not degrade the sorbent. The DOE authorized extending the project into the spring of 2015 to collect additional data from ADAsorb operation after the risk of freezing had passed, and pilot repairs were completed.

6.4.2 Second Round of Parametric Testing

Trend data of adsorber flow rate, sorbent flow rate, regenerator temperature, and CO_2 removed during continuous testing between 6/3/2015 and 6/25/2015 is shown in Figure 51. Operating conditions compared to design conditions are presented in Table 19.

The sorbent circulation rate was not measured directly due to challenges in measuring the flow of solids in a system. The sorbent circulation rate was calculated using the Darcy-Weisbach equation:

$$\Delta p = \lambda \left(l / d_h \right) \left(\rho v^2 / 2 \right) \qquad (13)$$

The Δp was measured using two pressure taps located at the outlet of the regenerator as sorbent is conveyed to the adsorber. The equation can be rearranged to yield density, and since the density of the sorbent is much greater than the gas, this value can be used along with the measured gas flow to determine the sorbent flowrate in lb/hr.

The bottom adsorber bed was the most critical level to control in the system. If the bottom bed level dropped below the minimum level, the difference in pressures between the adsorber and regenerator would cause a flow of flue gas into the rich sorbent transport line. When that occurred, a large decrease in the adsorber exit flow corresponded to a high regenerator outlet flow, typically resulting in significant sorbent carryover from the regenerator cyclone to the baghouse. Maintaining bottom bed level became the most critical operator control variable for stable operation. Increasing the sorbent circulation rate between the adsorber and the regenerator tended to cause the bottom bed level to drop, as sorbent would exit the bottom bed faster than it could be replenished. The highest continuous sorbent circulation rates tended to correspond to lower than design levels in the middle bed, which encouraged sorbent turbulence and filling of the middle to bottom bed drop tube. Opening the level control valve between the middle bed and bottom bed beyond approximately 50% open resulted in a reversal of flow through the drop tube, preventing sorbent flow between the middle bed and bottom bed of the adsorber. At times the reversal was substantial enough to cause the middle bed to fill beyond maximum level and starve the bottom bed of sorbent. In those situations, only a dramatic decrease in adsorber fluidization flow would break the stagnation and allow commencement of sorbent circulation.

The top adsorber bed level had a direct correlation to the level of sorbent in the lean sorbent cyclone dip leg (CYC-104). In order to promote flow out of the cyclone dip leg, and increase the sorbent circulation rate, the most effective sorbent levels were maintained at much lower than design (18 ft vs 26 ft design). When the adsorber was operated with a high level in the top bed, sorbent circulation rates decreased with a subsequent increase in the potential for sorbent carryover from the adsorber cyclone (CYC-101) to the baghouse.

Bed levels in the adsorber and regenerator are dependent on the fluidization velocities, the mass of sorbent in the bed, and the temperature of the bed. The total mass of sorbent in the system was increased beyond the design capacity to compensate for the lower velocities and bed levels being used. Evidence of this is shown in the higher than design pressure drop through the adsorber, as measured at the exit of the flue gas blower (BLW-101). The design blower outlet pressure was 5.9 psi, but operations maintained a pressure of 6-7 psi continuously.

During the 2015 testing between June 2 and June 25, the regenerator was maintained at nominally 120°C, the design operation conditions, but dropped below this point at times as indicated in the figure. The flow rate of flue gas through the adsorber was varied between 360 acfm and 1,600 acfm throughout the period. Recall that the design condition for the pilot was 1,808 acfm. The design condition for sorbent flow was 31,107 lb/hr. The circulation rate of sorbent in the system was varied between 10,000 and 26,000 lb/hr throughout the period. Due to changes in material handling characteristics of the sorbent, it was impossible to maintain sorbent flow above about 20,000 lb/hr.

The results indicate that greater than 90% CO_2 removal was achievable during some test conditions, particularly at low flue gas (adsorber) flow rates. At gas flows of nominally 800 acfm or above, the CO_2 removal was typically below 50%.

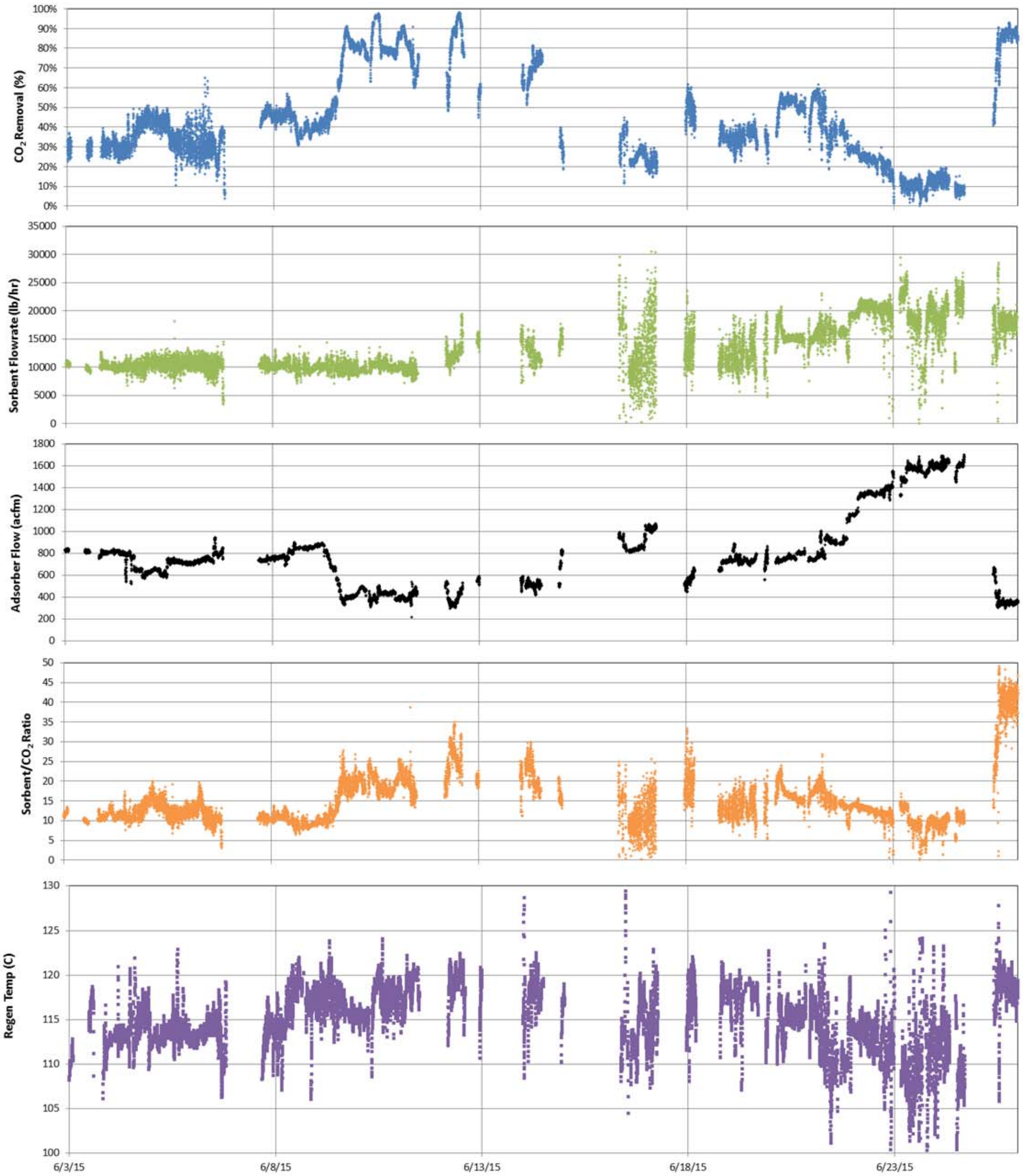


Figure 51. Overview of Data Collected after 6/6/2015.

Table 19. Pilot Design and Operating Conditions.

		Design	Measured						Units
Flue Gas	Adsorber In	1808	1611	779	824	428	415	350	acfm
Sorbent	Circulation Rate	31107	18439	15103	10070	13183	9876	17993	lb/hr
Sorbent/CO_2 Ratio	ADS Inlet	15.5	9.3	17.1	10.3	25.9	20.0	43.8	lb/lb
CO2 Inlet (%)	ADS Inlet	13%	13%	13%	13%	13%	13%	13%	%
CO_2 % Capture	% mass Captured	90.0%	13.6%	52.0%	42.0%	94.0%	82.0%	87.0%	Weight %
CO_2 Working Capacity	g CO_2/100g Sorbent	7.0%	1.5%	3.3%	4.4%	4.0%	4.3%	2.2%	gCO_2/100 g Sorbent
H_2O Working Capacity	g H_2O/100g Sorbent	1.0%		0.53%	0.86%	0.28%	0.27%		gH_2O/100 g Sorbent
CO_2	Fluidizing	343	262	267	195	237	229	306	acfm
	Transfer to REG	67	114	90	90	100	90	115	acfm
	Transfer to ADS	252	240	250	254	254	255	213	acfm
Density	Bottom ADS bed	15	15.3	19.8	19.3	20.7	22.4	23.3	lbs/cf
	Middle ADS bed	15	14.2	21.6	20.8	23.9	23.3	24.5	lbs/cf
	Top ADS bed	15	15.7	20.4	20	22.3	22.2	23.5	lbs/cf
	Regenerator	26	18	17.3	19	18.6	18.6	17.5	lbs/cf
Bed Height	Bottom ADS bed	13.3	11.5*	10.5	9	10.5*	7.5	7.8	ft
	Middle ADS bed	13.3	9.5*	7.5	8	8*	6	6	ft
	Top ADS bed	28.9	18*	17	21.5	22.5*	23	17.9	ft
	Regenerator	17	18*	17	16	20.5	15.5	15.9	ft
Temp	Bottom ADS bed	40	40	43	40	42	37	40	C
	Middle ADS bed	40	39	43	41	45	41	44	C
	Top ADS bed	40	43	51	42	43	41	46	C
	Regenerator	120	114	119	118	121	117	119	C
Adsorber In	BLW 101 Outlet	5.9	6.9	6	7	7	6.8	6	psig
	CO_2	13.1	13.1	13.7	13.8	13.8	13.1	14	%
Adsorber Out	ADS Outlet	1.0	1.43	0.38	0.45	0.32	0.32	0.3	psig
	CO_2	1.58	10.4	6.5	7.8	0.9	2.4	1.8	%
Regen Out	$CO_{2\,(dry)}$	93.4	89	96.8	85.5	88.8	85.5	89.9	$\%_{(dry)}$

* Unstable bed conditions during test

Data collected between 6/6/ and 6/24/15 was further analyzed to determine trends. The data set was scrubbed according to the conditions below for analysis.

- Calibration points generated during calibration of:
    - CO_2 Analyzer at adsorber inlet
    - CO_2 Analyzer at adsorber outlet
    - CO_2 Analyzer at regenerator outlet
- Data generated during:
    - Manual moisture measurement at CO_2 gas sampling locations
    - Filter blow-backs at CO_2 gas sampling locations
    - Regenerator temperatures less than 104°C
    - Other corrupt data as indicted by O_2 measurements in excess of 10% at CO_2 gas sampling locations

Percent CO_2 removal across the adsorber was calculated as follows:

$$\frac{\boldsymbol{Mass\ of\ CO_2\ in - Mass\ of\ CO_2\ out}}{\boldsymbol{Mass\ of\ CO_2\ in}} \tag{14}$$

A two-variable regression analysis was used to fit the entire scrubbed data set. The resulting model of CO_2 removal as a function of sorbent flow rate and adsorber flow rate is shown in Figure 52. The data fits this model with an R^2 value equal to 77% due in part to the fluctuations in operations due to material handling, including potential sorbent recirculation. The figure illustrates how at any given adsorber flow rate, an increase in sorbent flow rate produced an increase of CO_2 removal from the incoming flue gas. The most optimal conditions for CO_2 capture exist at low adsorber inlet flue gas flow rates and high sorbent flow rates. The operational data showed that using Sorbent BN within the as-built pilot, 90% CO_2 removal could be accomplished only at low adsorber flue gas flow rates combined with high sorbent circulation rates.

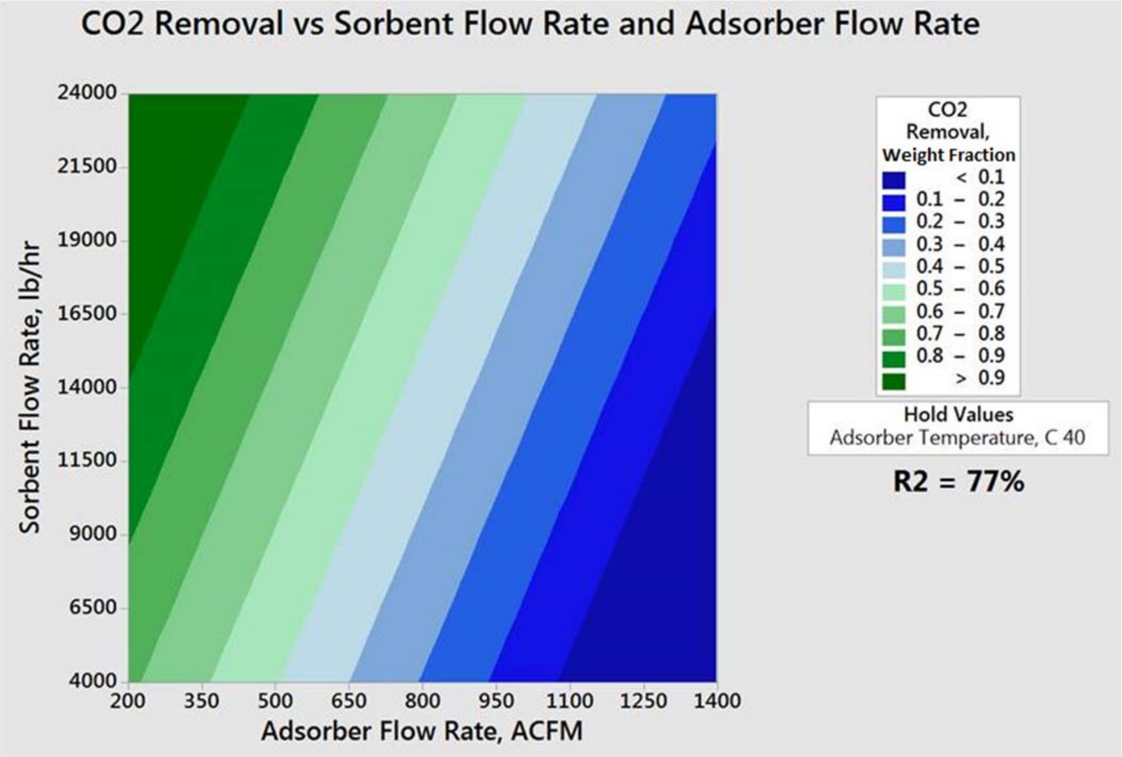


Figure 52. CO_2 Removal versus Sorbent Flow Rate.

A Pareto chart of the standardized effects determines the magnitude and the importance of an effect. The chart displays the absolute value of the effects and draws a reference line on the chart. Any effect that extends past this reference line is statistically significant. The Pareto chart shown in Figure 53 shows that the sorbent flow rate, and subsequent adsorber residence time, has the greatest impact on working capacity.

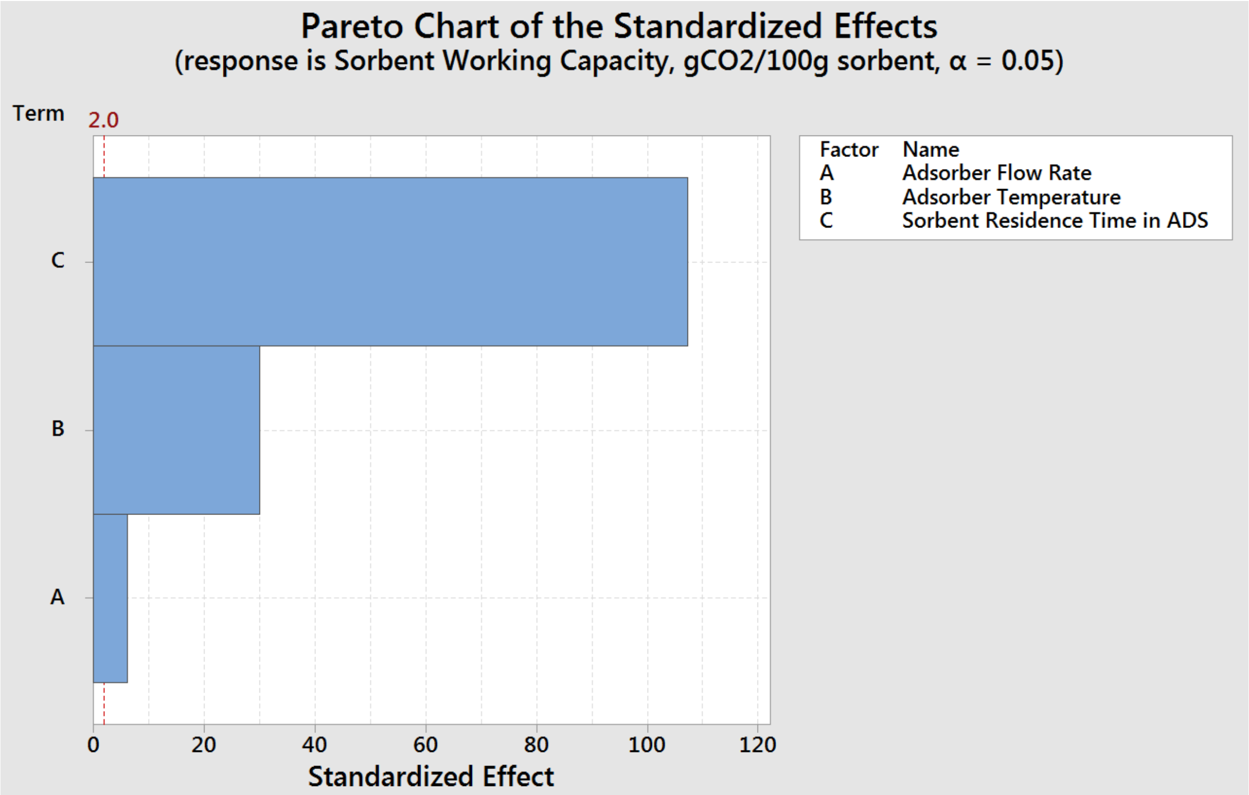


Figure 53. Pareto Chart Associated with Figure 52 CO_2 Removal Contour Plot.

Sorbent residence time within the adsorber is directly related to the sorbent flow rate combined with the sorbent density within the adsorber, which is dependent upon the adsorber inlet flue gas flow rate. A plot of sorbent residence time as it relates to sorbent flow rate can be seen in Figure 54.

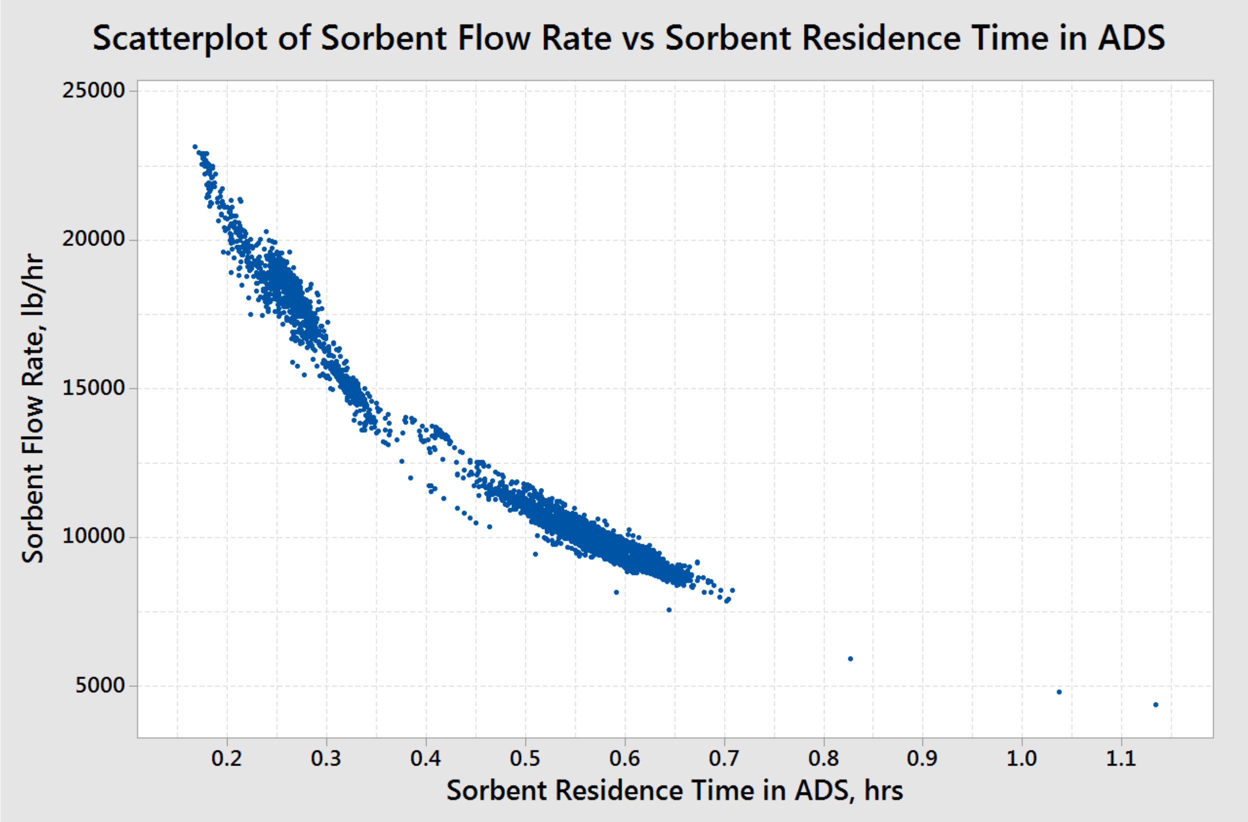


Figure 54. Plot of Sorbent Flow rate vs. Sorbent Residence Time in Adsorber.

A plot of the CO_2 capture compared to the ratio of the sorbent circulation rate/CO_2 mass flow for a range of flue gas flow rates through the adsorber is presented in Figure 55. For illustration, a line showing the CO_2 capture by a sorbent with a CO_2 delta loading across the adsorber of 5.4 g CO_2 /100g sorbent is included in the Figure. Note that because the units on the x-axis in the Figure are the sorbent to CO_2 mass ratio, 100% CO_2 capture is expected at a sorbent to CO_2 ratio of 18.5 at a delta loading of 5.4 4 g CO_2 /100g sorbent working capacity. Results indicating more than 5.4 g CO_2 /100g sorbent working capacity would fall to the left of the line shown in the graph. Thus, it is clear from the data shown that 5.4 g CO_2 /100g sorbent working capacity is the effective maximum delta CO_2 loading across the adsorber that Sorbent BN demonstrated during pilot operation.

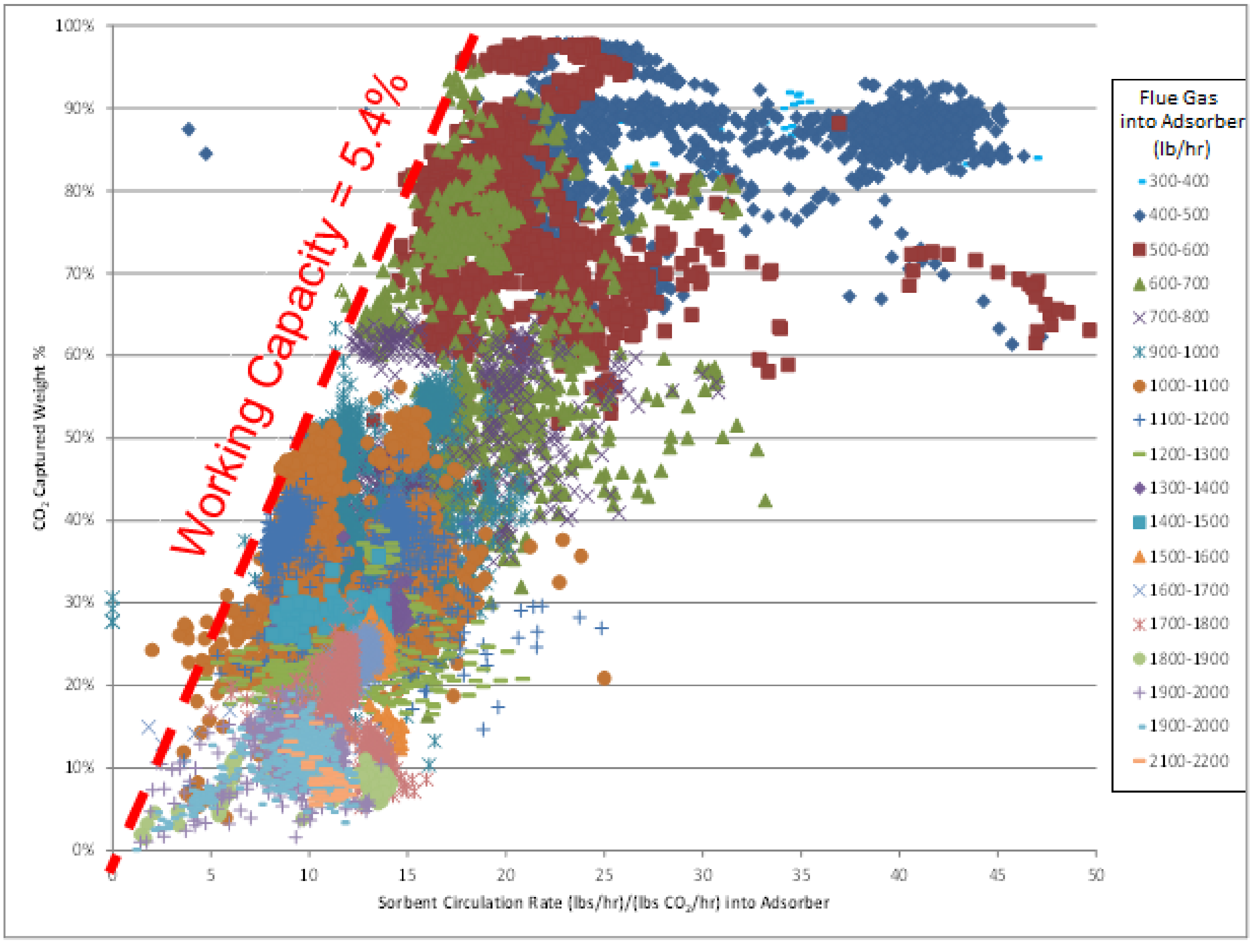


Figure 55. Plot of CO_2 % Removed versus Mass Ratio of Sorbent / CO_2 into Adsorber.

A review of pilot test results was conducted to determine key factors that were limiting the delta CO_2 loading on the sorbent across the adsorber and the CO_2 removal potential of the pilot. Potential reasons for delta CO_2 loading across the adsorber that were less than the working capacity expected from laboratory testing, as well as variability in the data to the right of the 5.4% line suggest other factors including the following.

- Degradation of sorbent in the pilot.
- Slower than expected reaction kinetics and corresponding insufficient sorbent residence time in the adsorber.
- Adsorption of CO_2 in the transport line between the regenerator and adsorber.
- Intermittent operation such as potential recirculation of sorbent to the regenerator when sorbent may have been bypassing the adsorber but recorded as passing through the adsorber, or periods with reduced regenerator temperatures.

6.4.3 Review of Potential Factors Impacting Performance during Parametric Testing

6.4.3.1 Sorbent Stability

The thermogravimetric analysis (TGA) performed on samples of sorbent taken from the pilot after Round 1 testing to assess the CO_2 capture performance indicated that no significant change in CO_2 capacity had resulted due to exposure to flue gas and system operation. Recall that during Round 1, the pilot experienced an upset condition when the temperature of the regenerator was initially increased to design temperature. The system was operated with flue gas for several days at 40°C to establish stable operation prior to this, however.

For this analysis, the sorbent was exposed to 100 % CO_2 at 120°C at first and subsequently CO_2 adsorption was assessed at temperatures in the range 120 to 40°C under 15 % CO_2 in N_2. Details of these results are shown in Figure 56.

Three sorbent samples were analyzed using the TGA to determine whether there was any sorbent degradation after exposure to flue gas. The samples were: BN13 (as-received), BN17 (sample taken in October from pilot) and BN 18 (sample taken in December from pilot). The results indicated that the samples collected during Round 1 of Parametric Testing did not show a substantial reduction in CO_2 working capacity between regeneration (120°C, 100 % CO_2) and adsorption (40°C, 15 % CO_2) conditions when compared to the as-received sorbent. Results of the particle size analyses also showed no change in particle size distribution from unused material.

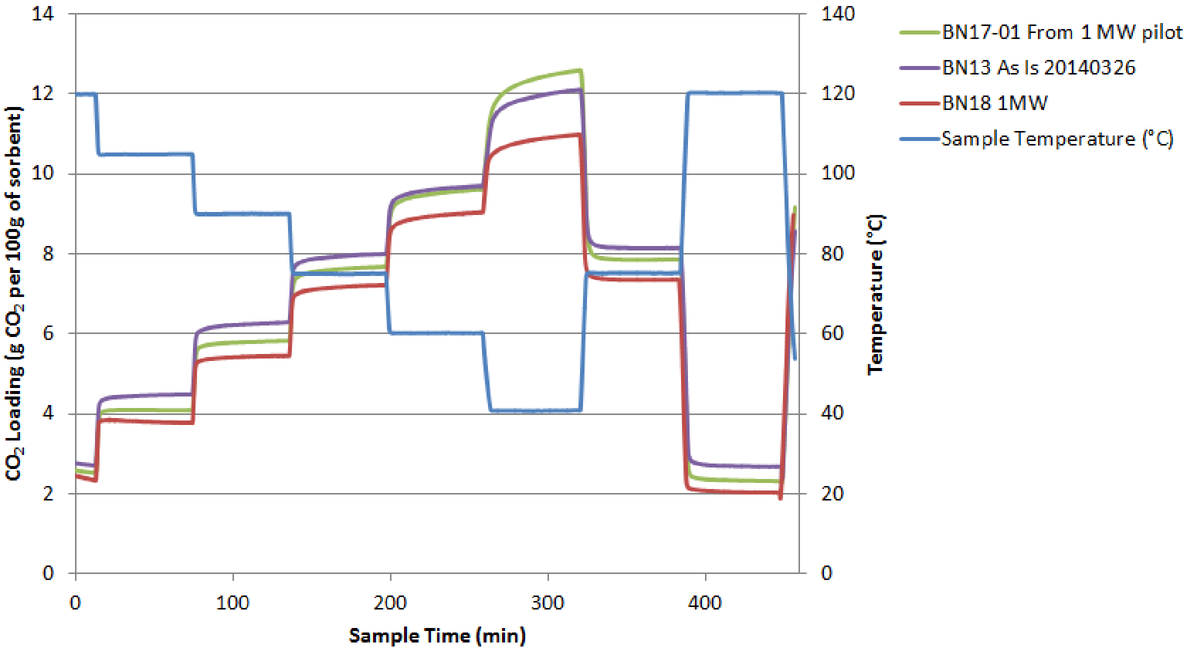


Figure 56. TGA Results for New and Used Sorbent Samples.

The thermogravimetric analysis (TGA) was again performed on sorbent taken from the pilot after 2015 testing to assess the CO_2 capture performance. This sample was taken after 23 days of continuous operation in the pilot, in addition to cycling in flue gas without regeneration for nominally 2 weeks during 2014 parametric testing. This TGA analysis also indicated that no significant change in CO_2 working capacity, when compared to sample BN 13. During the TGA test, the sorbent was again exposed to 100 % CO_2 at 120°C at first and subsequently CO_2 adsorption was assessed at temperatures in the range 120 to 40°C under 15 % CO_2 in N_2. The TGA weight versus time trend for the step-wise temperature test is shown in Figure 57.

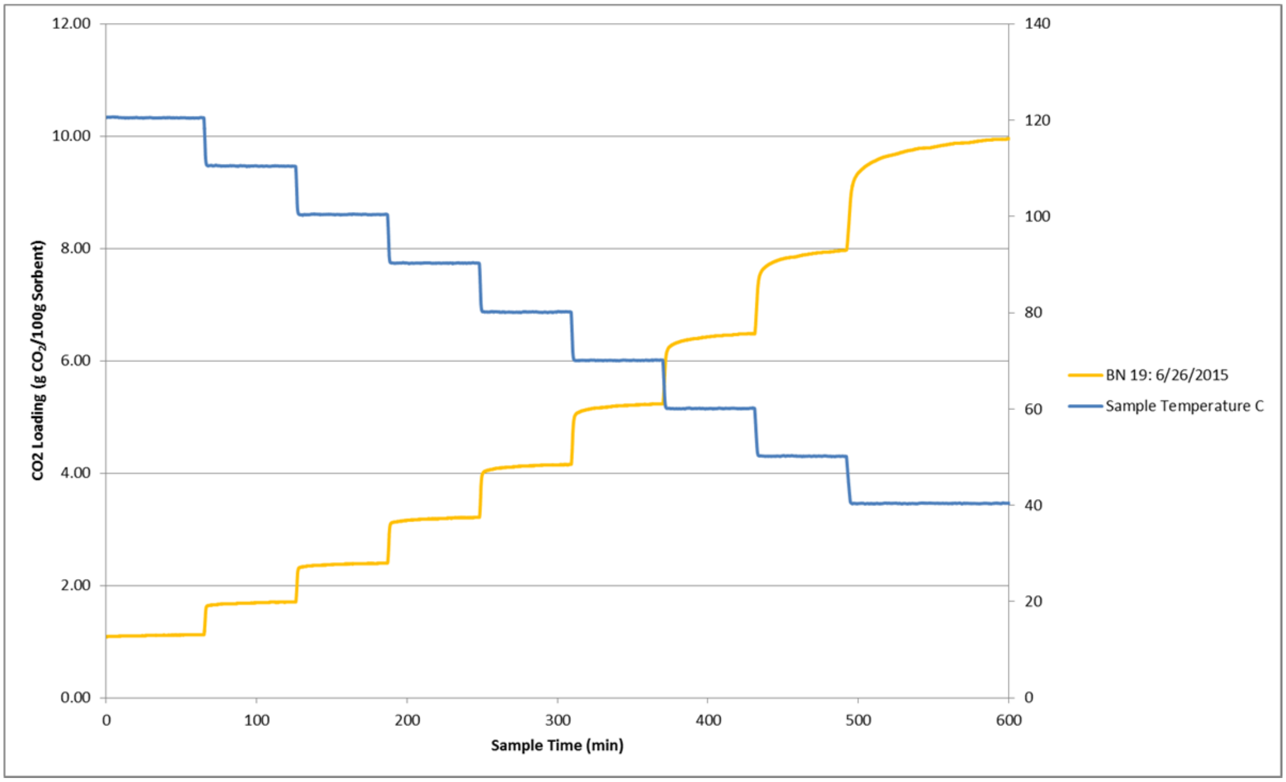


Figure 57. TGA Results for Sorbent Sampled from the Pilot on 6/26/15.

Laboratory data from TGA testing suggested that the equilibrium capacity of the sorbent under conditions present at the inlet and outlet of the adsorber in the pilot should approach 7%, as presented in Section 3.6.1. The CO_2 delta loading of the sorbent based on CO_2 measurements at the inlet and outlet of the adsorber compared to the mass flow of sorbent indicated that the working capacity within this portion of the system was 5.4 g CO_2/100g Sorbent (5.4%). Since laboratory data on sorbent removed from the system indicated that the equilibrium working capacity of the sorbent between the regenerator outlet conditions and the adsorber outlet conditions were relatively unchanged from sorbent prior to field testing, it is likely that there was another explanation for the limited delta CO_2 capacity calculated from pilot testing. Two potential explanations are non-equilibrium conditions at the exit of the adsorber, and loading the sorbent with CO_2, prior to entering the adsorber. These two potential causes are discussed further in subsequent sections.

6.4.3.2 Sorbent Reaction Kinetics

One potential cause of delta CO_2 loading below expectations was not reaching equilibrium loading because of insufficient residence time in the adsorber. Results from adsorption-desorption cycles in the TGA results and from the trickle-down reactor in the laboratory, presented in Section 3.6.2, indicated that the pilot design specification for adsorber residence time of 12 to 15 minutes should be sufficient to reach equilibrium loading. However, pilot results, as summarized in Figure 58, indicate that at least 30 minutes was required to reach 5.4 g CO_2/100g sorbent delta loading on the sorbent. Because additional residence time in the scrubber did not increase the delta loading above 5.4 g CO_2/100g sorbent, another factor likely contributed to both the limited CO_2 working capacity and could have contributed to the perception that sorbent kinetics were limiting pilot performance.

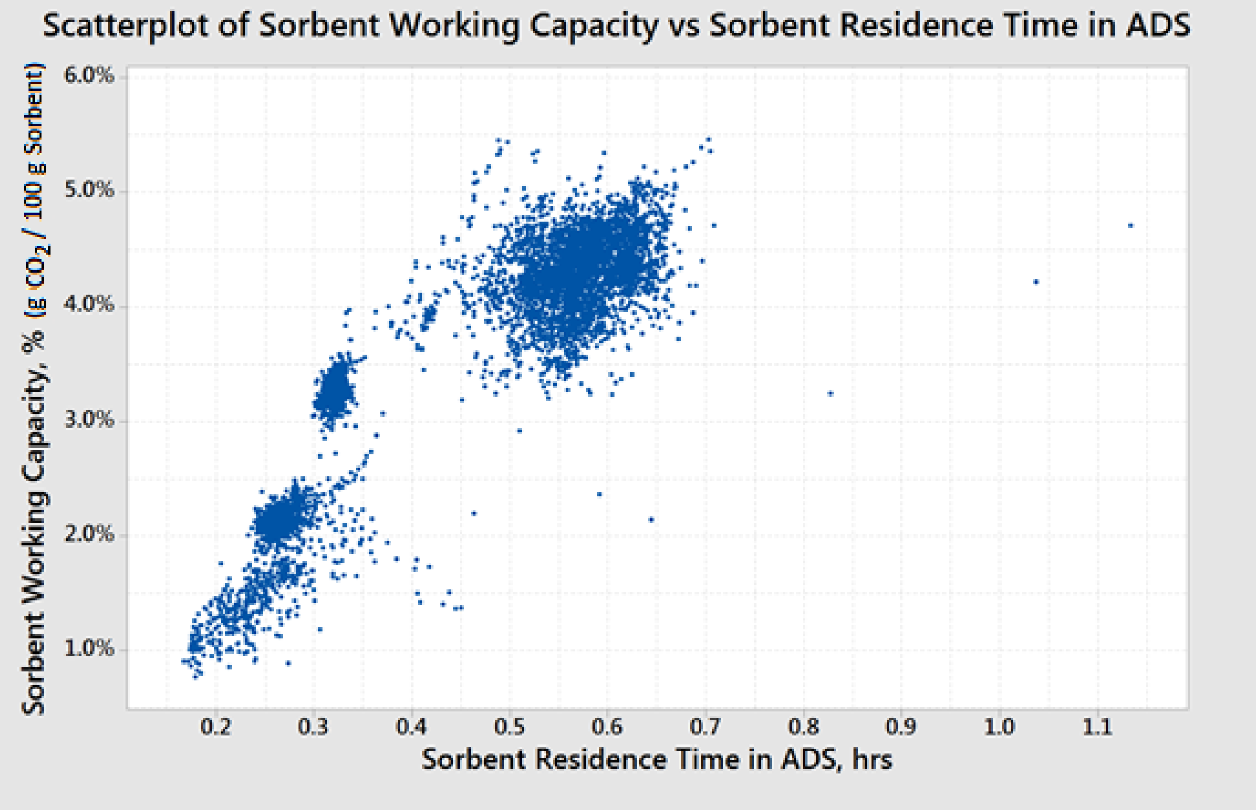


Figure 58. Plot of Sorbent Working Capacity versus Sorbent Residence Time in the Adsorber.

A statistical analysis of the relationship of sorbent working capacity to adsorber flue gas flow rate and adsorber residence time is illustrated in Figure 59. Working capacity is seen to be relatively unchanged by the rate at which flue gas is introduced into the adsorber where the residence time within the adsorber has a dramatic effect on the working capacity achieved. On average, residence times in excess of 40 minutes are shown to be needed within the adsorber in order to achieve the maximum observed working capacity of 5.4 g CO_2 /100 g sorbent.

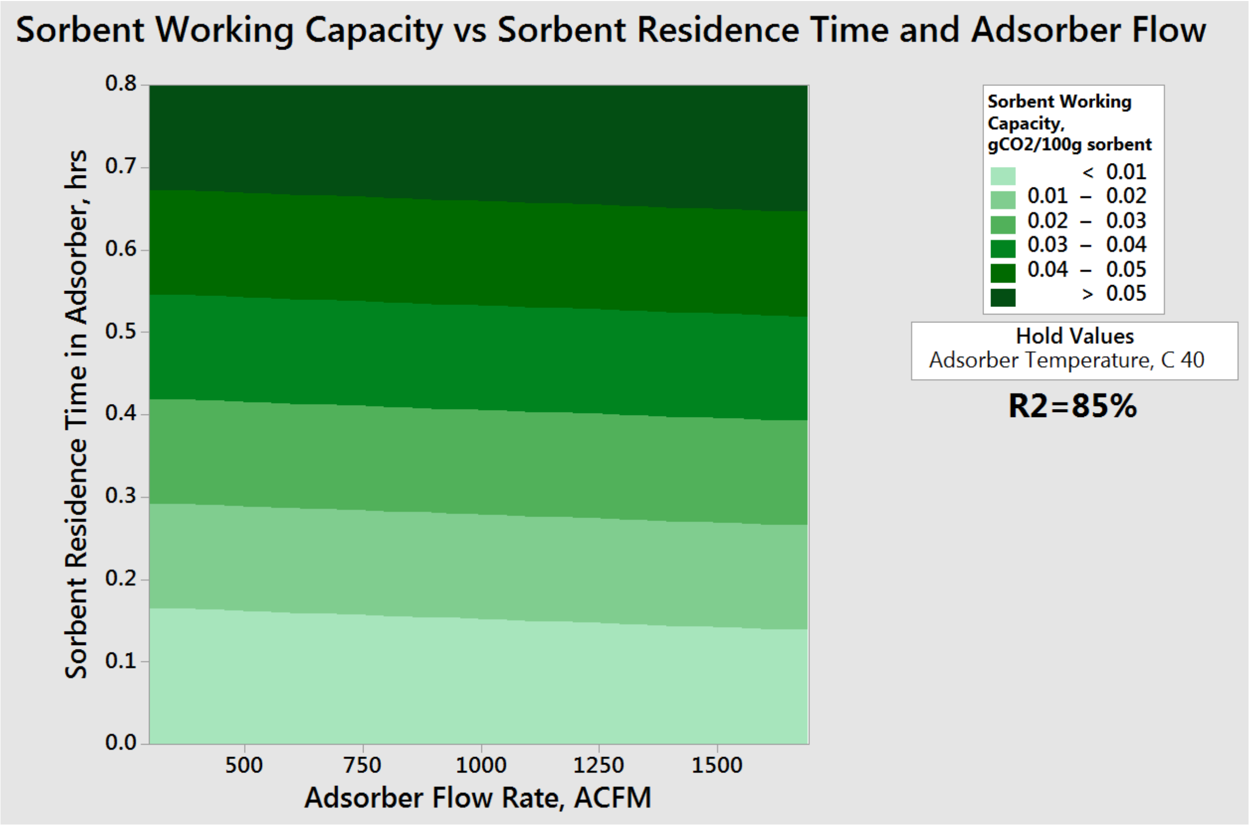


Figure 59. Sorbent Working Capacity versus Residence Time within the Adsorber.

6.4.3.3 Pre-Loading with CO_2 Prior to Adsorber

The most likely cause of the constrained working capacity is associated with the conveying gas used to transport sorbent from the regenerator to the adsorber. The conceptual design approach, as described in Section 4.1, was to use CO_2-lean flue gas to minimize adsorption of CO_2 onto regenerated sorbent. Because of the risk of oxidation of amines at elevated temperatures in the presence of oxygen, as described in Section 3.2.4, a last-minute design change was made prior to pilot testing to use oxygen-free gas. Rather than using nitrogen, which would have required adding a large nitrogen generator and other design changes in the pilot that may not be feasible at full scale, a decision was made to use CO_2-rich gas from the outlet of the regenerator blower (identified as BLW-103 in Figure 60) as the conveying gas. Although the risk of CO_2 uptake was noted, the risk was initially discounted because of the misconception that the nominal two seconds of residence time during sorbent conveying was not sufficient to result in significant CO_2 adsorption in the transport line.

The regenerator blower outlet temperature at the regenerator fluidizing and carrier lift gas location was nominally 75°C, well below the regeneration temperature of 120 °C. The cooled blower exit gas was used to transport the regenerated sorbent from the regenerator outlet to the adsorber. The average carrier gas flow was 290 scfm. The stream was 80% to 90% CO_2, which results in nominally 1,800 lb/hr CO_2 in the carrier gas. Based on the TGA analysis, at 75°C and a 0.8 bar partial pressure (see Figure 15), the sorbent loading is at equilibrium around 8.5 g CO_2/100 g sorbent (note that this is total loading, not working capacity). Results from Run 8 of the trickle-down reactor test in the laboratory (Table 7) indicated that less than two seconds was sufficient for sorbent BN to adsorb 2.6g CO_2/100g sorbent at 40°C from a gas stream with 12.5% CO_2, which is far enough from equilibrium capacity that sorbent kinetics are expected to be very fast based on both TGA testing and the trickle-down reactor tests. Additionally, a thermocouple that was installed in the dip leg below CYC-104, the cyclone designed to separate sorbent from the carrier gas, indicated that the sorbent had cooled to 63°C prior to entering the adsorber for that manual measurement. It is very plausible that sorbent BN adsorbed additional CO_2 while being conveyed between the regenerator and adsorber. Since the loading exiting the regenerator is expected to be nominally 3.45 lb CO_2/100 g sorbent at 120°C and pilot regenerator pressure, and the equilibrium loading is expected to be 10.47 at 40°C and the adsorber inlet operating pressure, only 1.62 additional g CO_2/100g sorbent would need to be adsorbed in the transport line to limited the delta loading in the adsorber to 5.4 g CO_2/100g sorbent. Thus, the overall working capacity <u>between the regenerator outlet and adsorber outlet</u> would be 7.0 g CO_2/100g sorbent (5.4 g CO_2/100g sorbent from the adsorber and 1.6 g CO_2/100g sorbent from the transport line), as expected from laboratory tests.

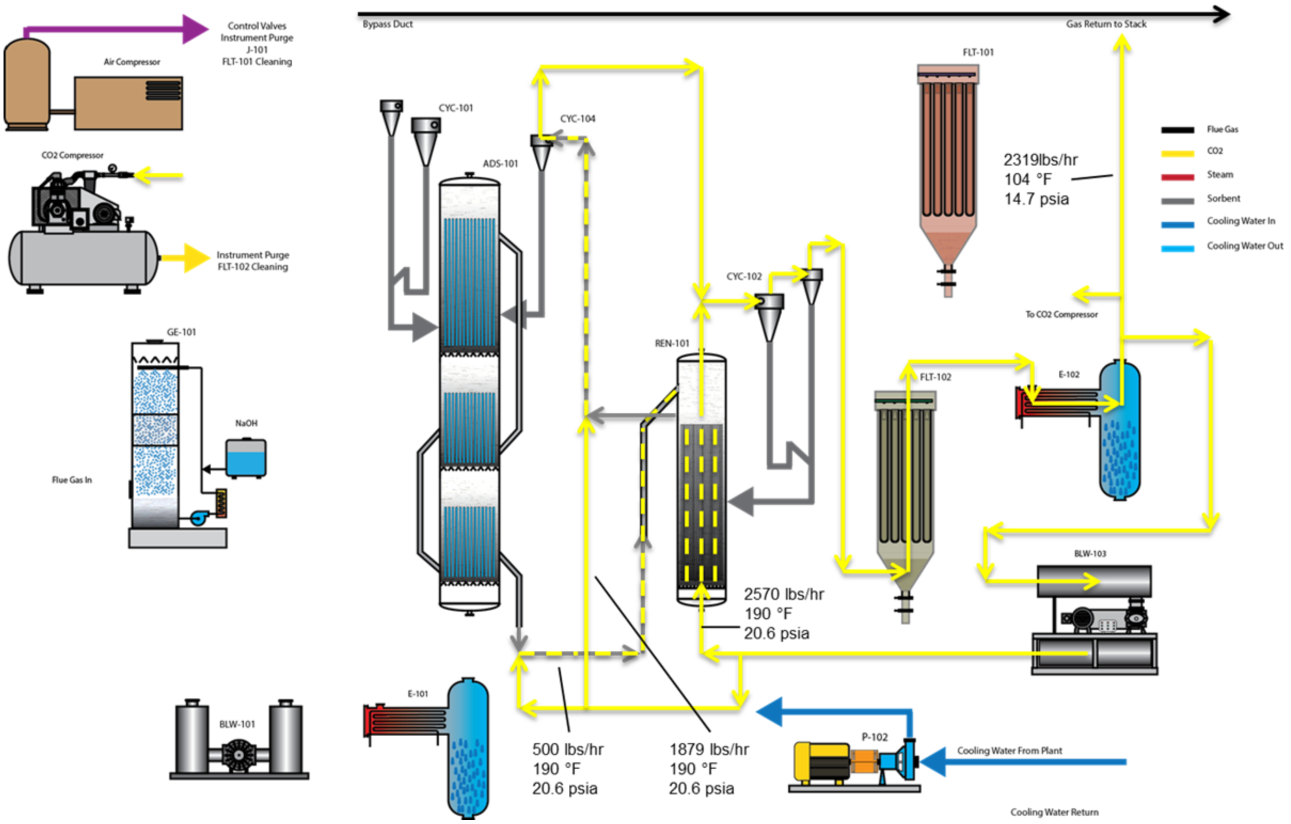


Figure 60. CO_2 flow during ADAsorb Pilot Testing.

The impact of sorbent flow on CO_2 removal is further illustrated by Figure 61 and Figure 62. In both Figures, the data set is divided into two groups by color, sorbent flow rates below 12,000 lb/hr in light blue, and sorbent flow rates above 12,000 lb/hr in dark blue. Figure 61 shows the variability of CO_2 removal over the range on sorbent flow rates tested. Figure 62 indicates that, in general, the results where the adsorber delta loading approached 5.4 g CO_2/100g sorbent coincided with lower sorbent flow rate. Although high CO_2 removal could be achieved with high sorbent circulation rates, increasing the circulation rate resulted in lower delta CO_2.

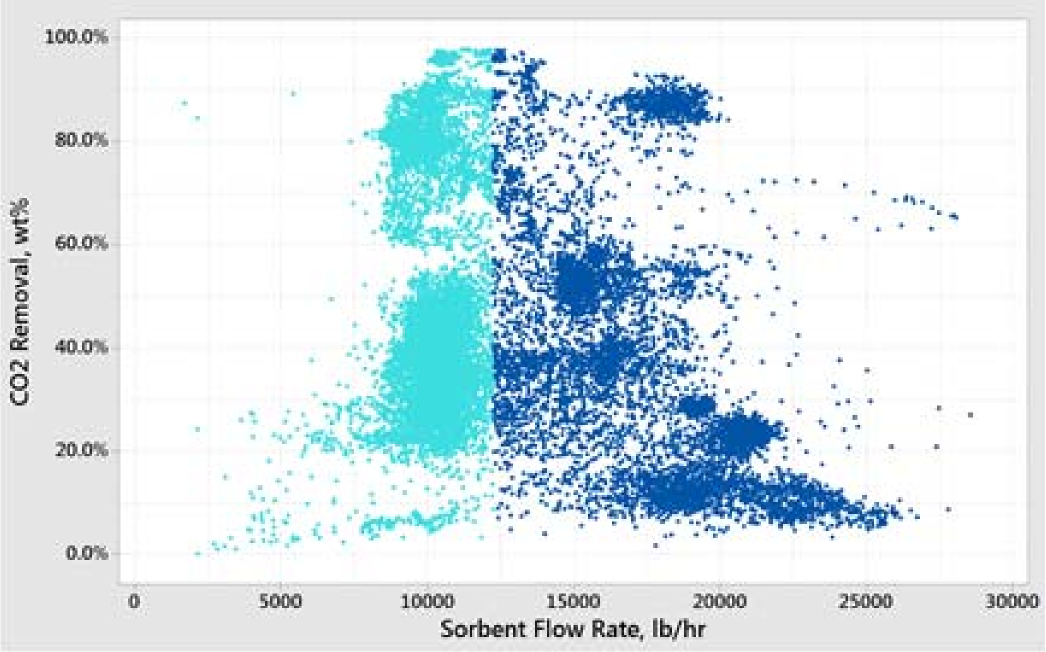


Figure 61. Plot of CO_2 Removal versus Sorbent Flow Rate

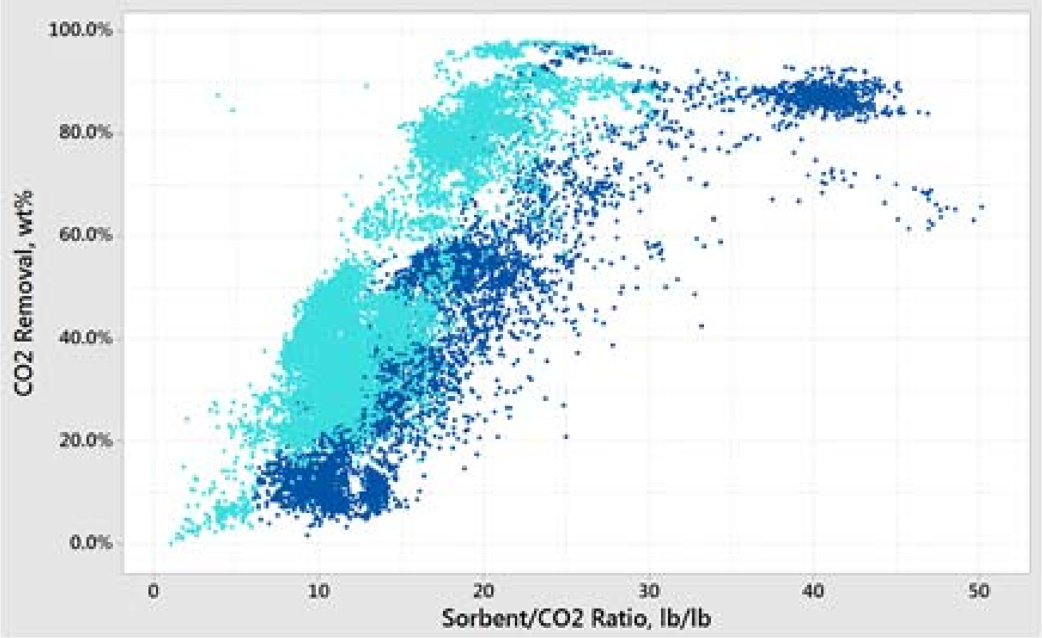


Figure 62. Plot of CO_2 Removal versus Sorbent/CO_2 Weight Ratio; colors indicate sorbent flow rate as shown in Figure 61.

A model was developed to better understand the potential impact of CO_2 adsorption in the transport line between the regenerator and adsorber, and the resulting impacts on pilot performance. A sketch of the 3-stage reactor showing key model inputs (in green) and select calculated parameters, is shown in Figure 63. The simplified model assumed isothermal conditions in the adsorber at 40°C, and system pressures that were consistent with both the design conditions and operation. A curve fit of the 40°C sorbent isotherms from laboratory testing for the range of partial pressure expected in the adsorber was incorporated into the model, where

$$CO_2 \text{ loading} = 2.4565\ln(P,\text{bar}) + 15.557,\ R^2 = 0.9923 \qquad (15)$$

The flue gas flow, inlet CO_2 concentration, and sorbent recirculation rate were independent variables in the model. Data from pilot testing was used to calibrate the model for the amount of CO_2 that was entering the adsorber with the return sorbent. As noted previously in this section, the equilibrium loading based on the TGA analysis at the conditions in the transport line between the regenerator and the adsorber were nominally 75°C and 80% to 90% CO_2. At 75°C and 0.8 bar partial pressure is 8.5 g CO_2/100 g sorbent (note that this is total loading, not working capacity).

An output of the model compared to pilot results is presented in Table 20. As shown in the table, if the sorbent loading entering the adsorber is consistent for all conditions at 6.3 g CO_2/100 g sorbent, which is below the equilibrium loading for conditions in the transport line, the model output for CO_2 removal across the adsorber is in good agreement with actual pilot test results (see cases 2 through 4 in the table). For case 1, the sorbent circulation rate was near the limit of stable pilot operations. The low measured CO_2 removal in the pilot compared to model results suggests that a portion of the sorbent was recirculating back into the regenerator through cyclone 104, as discussed in Section 6.3.2.

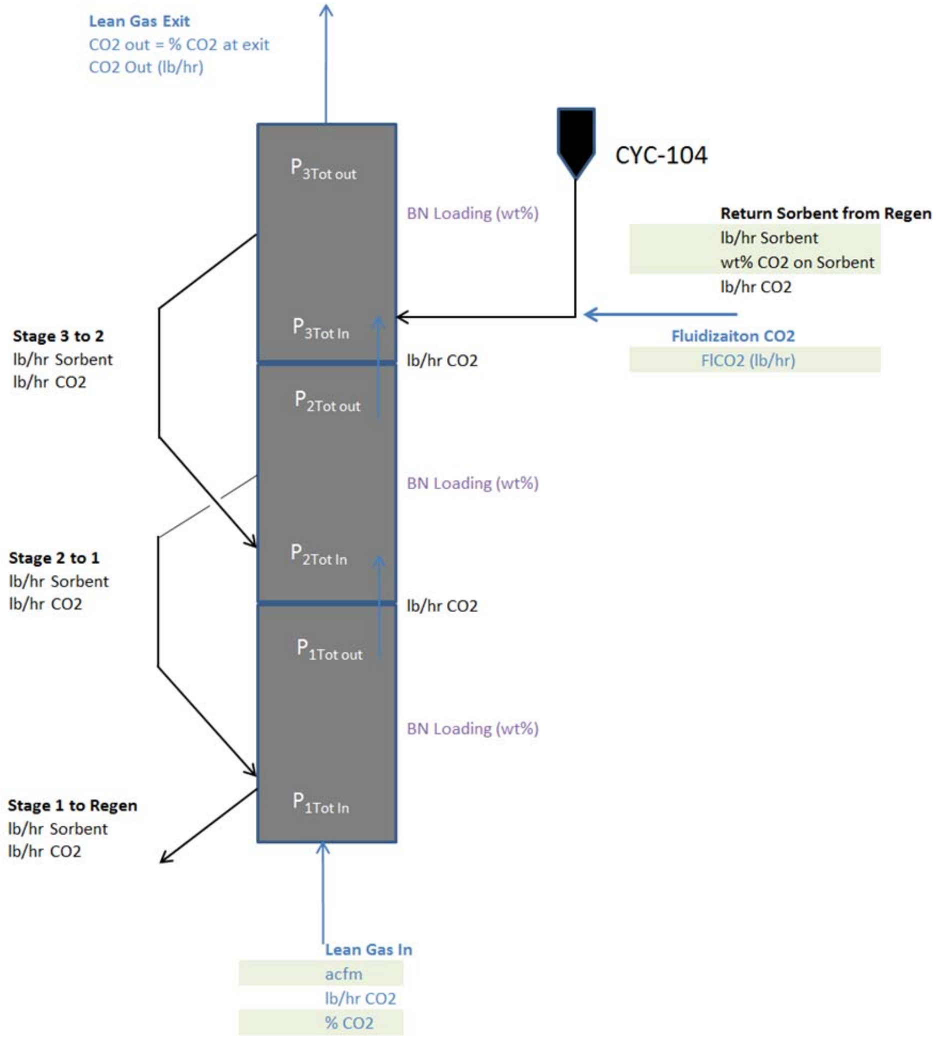
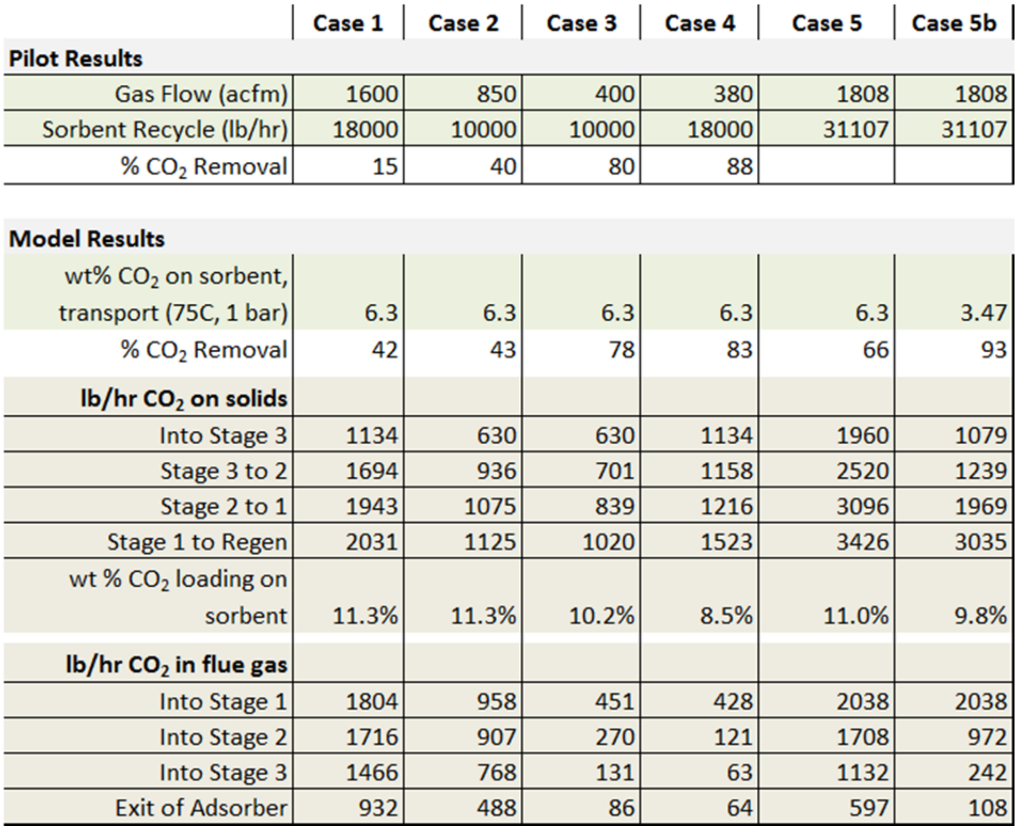


Figure 63. Pilot Model Overview.

Two additional model runs are shown in Table 20. Case 5 shows the pilot design conditions for adsorber flow rate and sorbent circulation rate with the additional CO_2 added to the transport gas. Case 5b shows the design conditions without additional CO_2 collected in the transport line. As shown, greater than 90% CO_2 removal is expected at the pilot design conditions with sorbent BN if the CO_2 addition in the transport line is eliminated. These data are shown with the pilot results graphically in Figure 64. Recall that there were also handling issues associated with sorbent BN at elevated temperatures that would require physical modifications to the cyclone and dip leg at the outlet of the regenerator to accommodate issues associated with handling hot sorbent.

Complete characterization of each bed in the adsorber using field measurements was not possible due to sampling issues. Occasional spot checks could be performed with a portable CO_2 analyzer within the disengaging space between the bottom and middle beds and between the middle and top beds. These spot checks indicated that the majority, up to 80%, of the capture occurred in the top bed; however, the data is questionable as the analyzers sampled through an extraction system which could have had a pressure slightly higher than ambient and which did not cool the gas sample. Given the questionable nature of these readings and the fact that they were not continuous, it was not practical to determine a mass balance around each bed from actual measurements. The model developed does provide insights, however. When the sorbent recirculation rate compared to CO_2 entering the system is low, more CO_2 is captured in the top bed because the sorbent approaches capacity before cascading into lower beds. As the sorbent/CO_2 ratio increases, the fraction of CO_2 captured in the top bed approached zero because the CO_2 captured in the gas is adsorbed by the excessive sorbent prior to reaching the top bed. Furthermore, the model suggests that as the CO_2 removal increases, the benefit of three stages versus two stages declines because most of the removal is occurring in the lower two beds. These data are shown in Figure 65 for both the case with CO_2 in the transport gas from the regenerator to the adsorber contributing to the loading on the sorbent, and for the case where the no additional CO_2 from the transport gas is adsorbed by the sorbent.

Table 20. Pilot Model Output.



	Case 1	Case 2	Case 3	Case 4	Case 5	Case 5b
Pilot Results						
Gas Flow (acfm)	1600	850	400	380	1808	1808
Sorbent Recycle (lb/hr)	18000	10000	10000	18000	31107	31107
% CO_2 Removal	15	40	80	88		
Model Results						
wt% CO_2 on sorbent, transport (75C, 1 bar)	6.3	6.3	6.3	6.3	6.3	3.47
% CO_2 Removal	42	43	78	83	66	93
lb/hr CO_2 on solids						
Into Stage 3	1134	630	630	1134	1960	1079
Stage 3 to 2	1694	936	701	1158	2520	1239
Stage 2 to 1	1943	1075	839	1216	3096	1969
Stage 1 to Regen	2031	1125	1020	1523	3426	3035
wt % CO_2 loading on sorbent	11.3%	11.3%	10.2%	8.5%	11.0%	9.8%
lb/hr CO_2 in flue gas						
Into Stage 1	1804	958	451	428	2038	2038
Into Stage 2	1716	907	270	121	1708	972
Into Stage 3	1466	768	131	63	1132	242
Exit of Adsorber	932	488	86	64	597	108

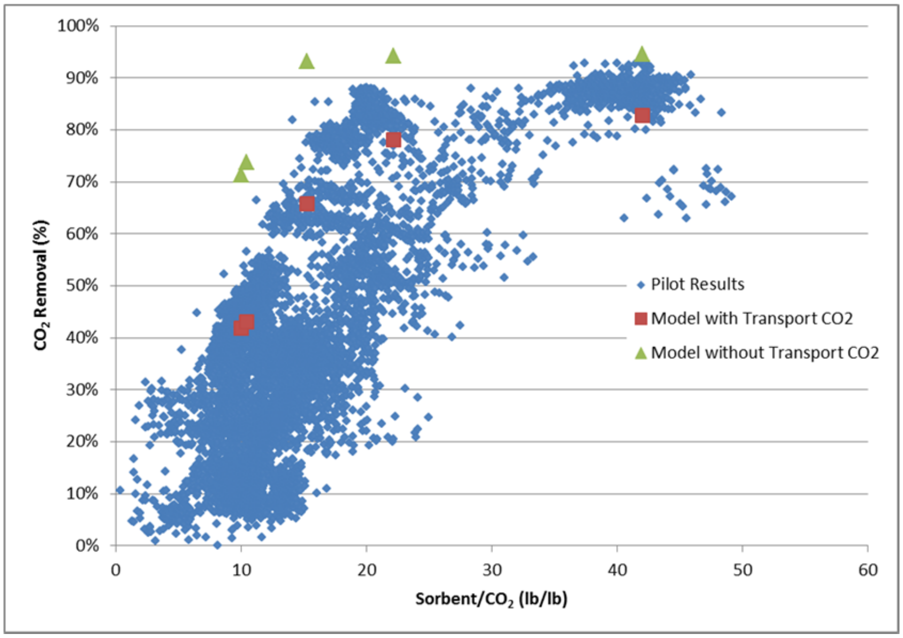


Figure 64. Model Results Compared to Pilot Results, with and without CO_2 Adsorption in the Transport Line.

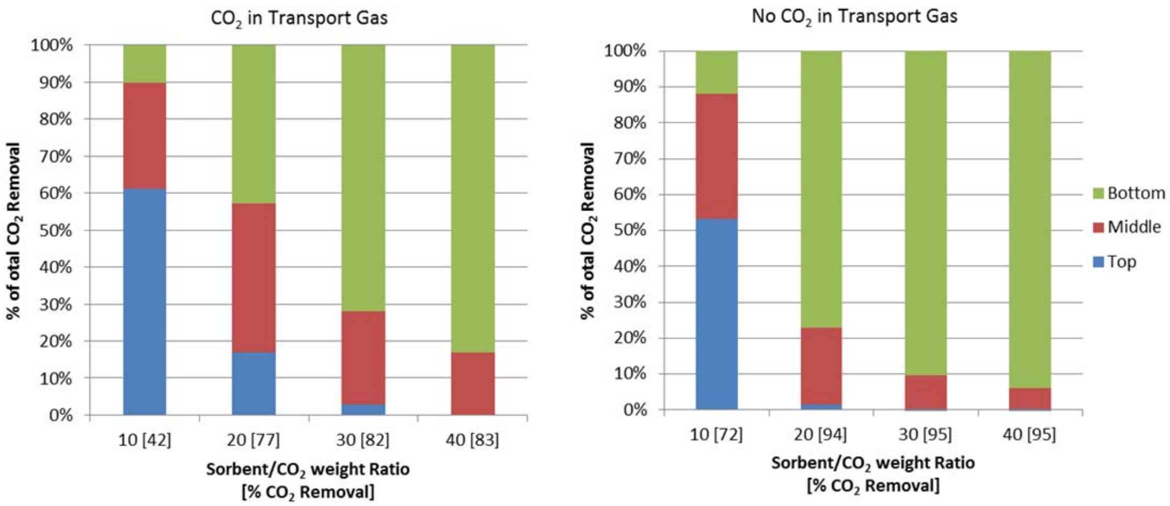


Figure 65. Model Results: Fraction of CO_2 removed in Each Stage, with and without CO_2 Adsorption in the Transport Line.

Additional analysis was conducted using the model in conjunction with the pilot data to determine the influence of sorbent residence time in the adsorber, or sorbent kinetics, on sorbent adsorption capacity.

From Figure 58, it appeared as if sorbent kinetics was a significant factor in the overall performance of the sorbent. However, when including the impact of CO_2 entering stage 3 of the adsorber as a function of sorbent recirculation rate, slow adsorption kinetics do not appear to be a factor in pilot performance, as indicated in Figure 66.

Using heated CO_2 to convey sorbent from the regenerator to the adsorber should mitigate adsorption in the conveying line. A short test was conducted early in the pilot testing effort to determine whether heating the carrier gas would result in improved performance. Unfortunately, this was done during the period when signal grounding issues were being experienced. No change in working capacity was noted during this test and the test was not repeated during the final parametric testing period after many of the operational issues had been resolved.

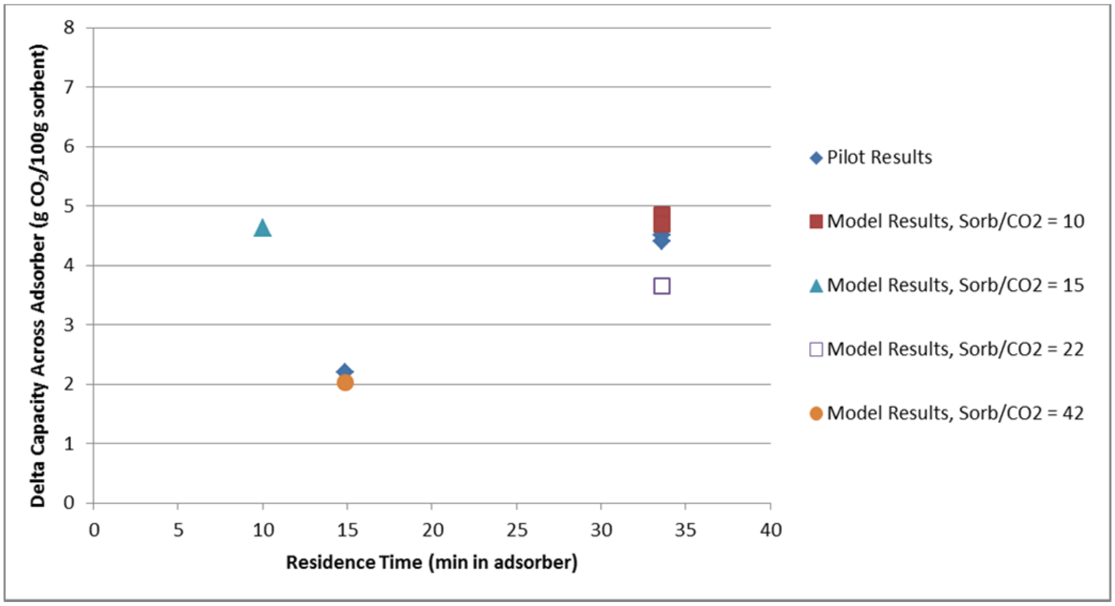


Figure 66. Model Results of Sorbent Capacity as a Function of Sorbent Residence Time in Adsorber and Sorbent/CO_2 Flow Rate.

An alternate theory, suggested through modeling efforts by the CCSI team, was insufficient regeneration of sorbent due to non-isothermal conditions predicted in the regenerator could be a limiting factor on overall CO_2 removal. A trend graph of regenerator temperatures is presented in Figure 67. Although it was not possible to maintain the regenerator temperature at 120°C throughout parametric testing, data on June 25, 2015 during the 20-hour test at stable conditions indicates that it was possible to maintain the bed temperature within 10°C of the set point, which is much better than that predicted by the CCSI modeling team. However, the sorbent temperature most representative of the gas exiting the regenerator is the thermocouple identified as "Mid Discharge" in the figure, which was nominally 5°C below the set point temperature during the 20-hour continuous test. This temperature difference could account for approximately 1 g CO_2/100 g sorbent being retained on the sorbent compared to the 120°C equilibrium loading, which is 14% of the expected delta loading of 7 g CO_2/100g sorbent. Since the operating model described in the section was only for the adsorber, there is no effect on the results of the operating model whether the CO_2 is added in the transport line or retained from the regenerator. Therefore, both of these factors should be considered when assessing options to improve pilot performance.

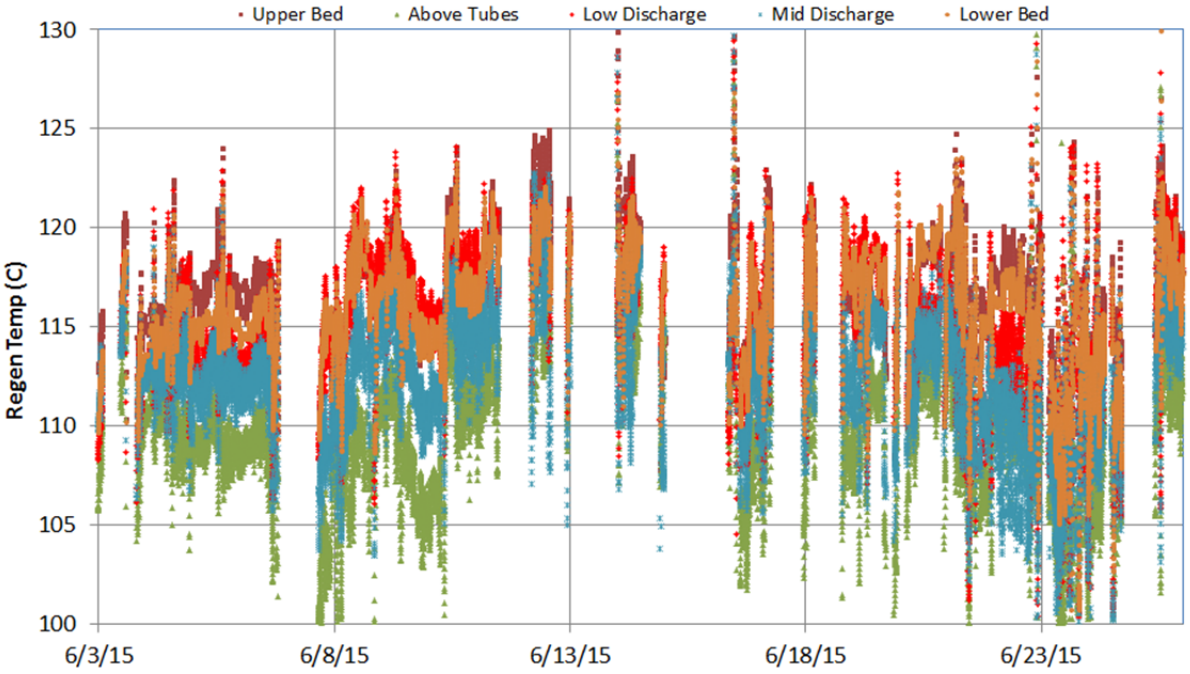


Figure 67. Variation in Regenerator Temperature During Testing.

6.4.3.4 Intermittent Operating Conditions

Most of the data analysis was conducted on the entire set of parametric data from round 2 of testing, with the caveats described at the beginning of this section. Some of the variability in the data set was from intermittent conditions. Many of these are discussed in this section.

The temperatures within each bed were measured at several points along the vertical axes to maintain adsorption conditions approaching isothermal. The measurements from the thermocouples in the top stage are shown for round 2 of parametric testing in Figure 68, with T-26 at the bottom of the bed, and T-31 at the exit of the bed. The isothermal conditions for testing were 40°C. It is clear from the trend graph that the bed was warmer at the bottom where sorbent from the regenerator was being added, and cooler at the top. It is also clear that it was difficult to maintain the temperatures much closer than within about 5°C of the set point in this dynamic system.

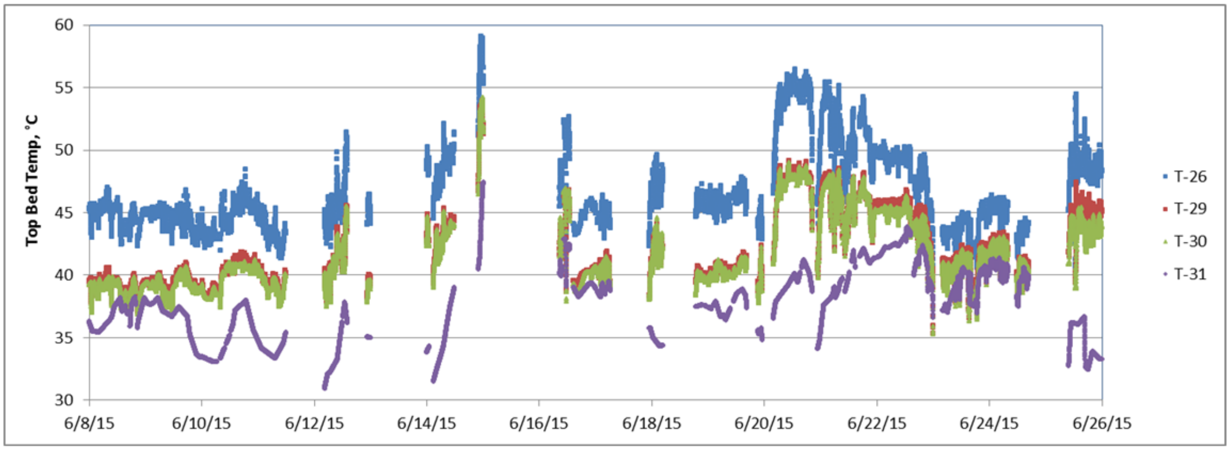


Figure 68. Temperature of Top Adsorber Bed during Parametric Testing.

In an effort to assess the effect of regenerator temperature on the operating system, two additional regenerator temperatures, 110°C and 115°C, were tested. The regenerator test was conducted on 6/11/2015, the day prior to the fluidizing N_2 and CO_2 campaign discussed earlier in this section. The assumption has been made that the rate at which the sorbent was potentially bypassing the adsorber was constant during the regenerator temperature testing time period due to the relatively short, six-hour test. Table 21 indicates that as the regenerator temperature increased, the sorbent working capacity increased, resulting in increased removal within the adsorber. This is due to increased removal of CO_2 from the sorbent within the regenerator leading to increased working capacity. Figure 69 and Figure 70 graphically display the change in working capacity and CO_2 removal as the regenerator temperature changes.

Table 21. Effect of Regenerator Temperature on Adsorber CO_2 Removal.

Regenerator Temp	Sorbent Working Capacity	CO_2 Removal	Sorbent Circulation	Flue Gas
°C	g CO_2/100g Sorbent	% mass	lb/hr	acfm
110.5	2.01	46%	10,893	457
115	2.15	48%	10,476	458
119.5	2.46	53%	10,141	458

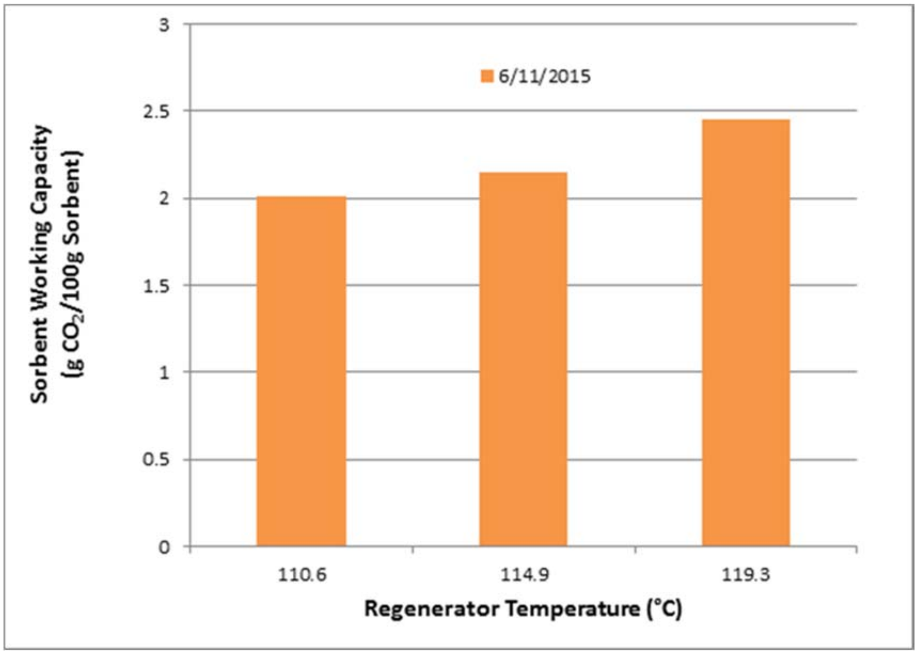


Figure 69. Plot of Sorbent Working Capacity versus Regenerator Temperature.

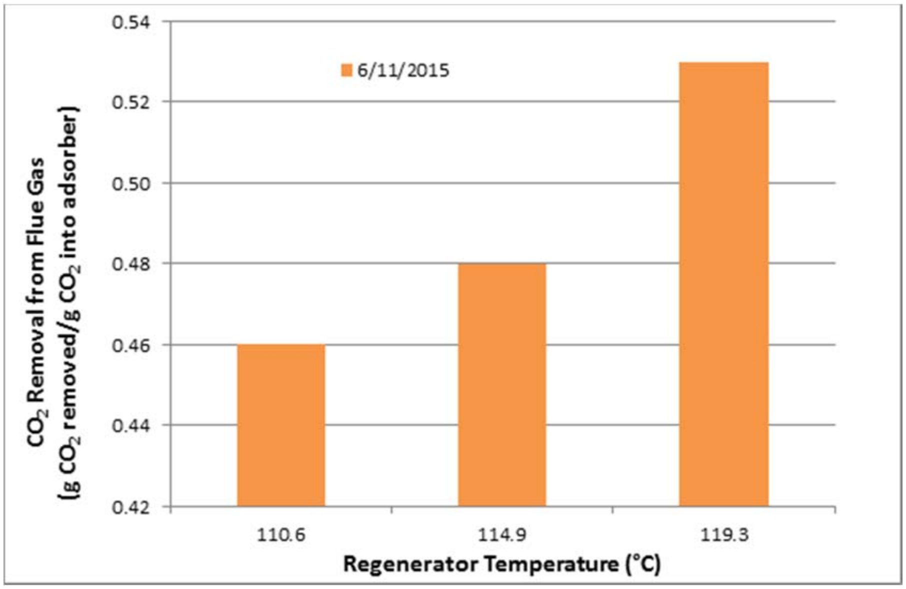


Figure 70. Plot of CO_2 Removal versus Regenerator Temperature.

As previously discussed, the sorbent circulation rate had the largest impact on the removal of CO_2 from the adsorber inlet flue gas stream. A statistical analysis of regenerator temperature and impacts on CO_2 removal is shown in Figure 70. The regeneration temperature also has a large impact for relatively low changes in temperature. Note that the regeneration temperature of 120°C is the upper limit for regeneration of sorbent BN due to higher temperature regeneration degrading the sorbent over time, as discussed in Section 3.6.2, and the results presented in Figure 69 range between 110 and 119°C.

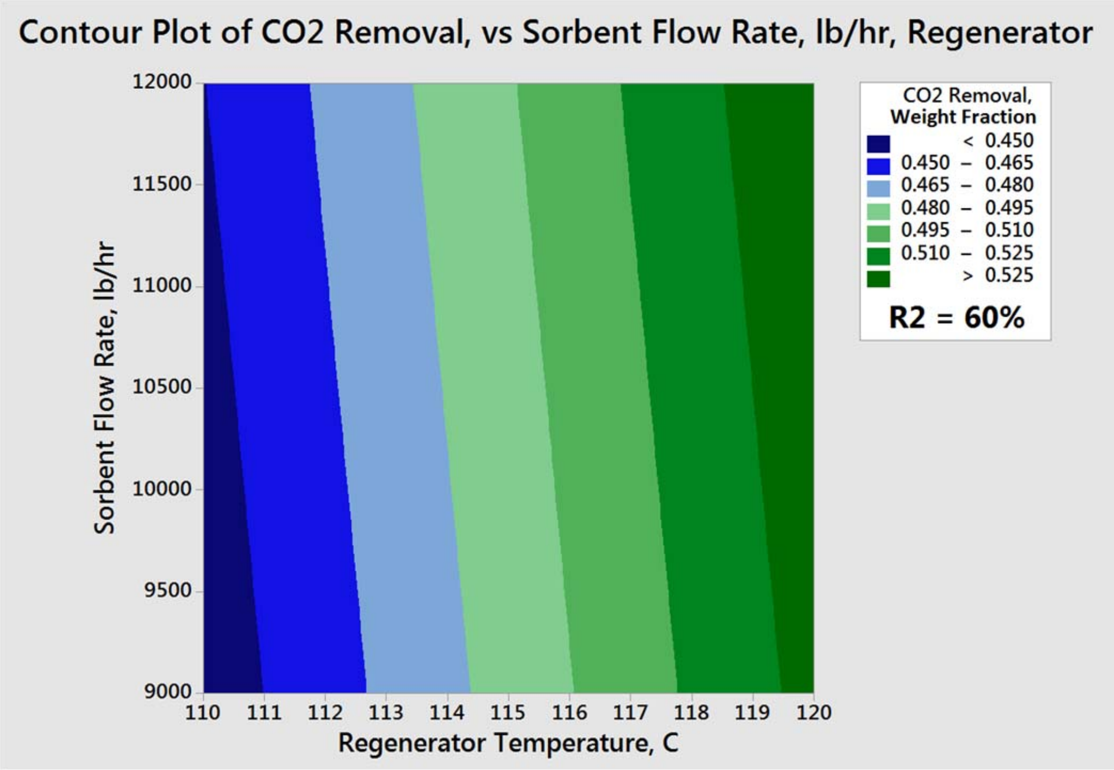


Figure 71. Effects of Regeneration Temperature and Sorbent Circulation Rate on CO_2 Removal.

The regenerator was designed and constructed with thermocouples located throughout the heat transfer surface and the free space above the tubes. It was observed that the temperature in the topmost location was slightly lower than the regenerator temperature, but that the total mass of sorbent in the regenerator was great enough, and close enough to the 120 °C equilibrium temperature to not experience the cooling effect as dramatically as predicted in the NETL CCSI study, as discussed in Section 3.7. Part of the effect could be attributed to sorbent circulation rates that were measured lower than design. Lower sorbent circulation allowed longer regenerator residence time, as well as less differential mass of incoming sorbent cooling the regenerator bed.

A number of times during operation, the regenerator bed slumped, or dropped to the bottom of the regenerator due to an equipment trip. During restart, significant difficulty was experienced in attempting to restart fluidization to the regenerator. The entrained moisture and subsequent cooling of the sorbent following a trip caused the sorbent to form concretions. Rapid flow fluctuation (pressure pulsation), and draining and reloading of sorbent was required to break loose the solidified sorbent.

Had the design incorporated direct injection of steam, as recommended in the NETL CCSI modeling report, it is likely that the solidification of sorbent would have been much worse, possibly ceasing all operations.

6.4.3.5 Moisture Uptake during Pilot Operation

Also of note is the measured moisture working capacity of the sorbent at each condition. The design value of 1.0 g H_2O/100 g sorbent was determined through experimental testing in a nitrogen atmosphere. Limitations of operation of the TGA did not allow for testing of selectivity between CO_2 and moisture. Pilot results from manual measurements show the average moisture uptake to be only 0.49 g H_2O/100 g sorbent, substantially lower than the design point. Manual measurement results are presented in Section 3.6.4.

6.4.3.6 CO_2 Product Purity

The CO_2 product purity from the pilot averaged 89.4% on a dry basis during 90% CO_2 removal from the flue gas.

Table 22. CO_2 Product Purity

		Design	Measured Values						
Operational Values	Flue Gas Flow Rate	1808	1611	779	824	428	415	350	acfm
	Sorbent Circulation Rate	31107	18439	15103	10070	13183	9876	17993	lb/hr
	CO_2 Removed witin Adsorber	2172	267	496	439	521	423	390	lbs/hr
Regenerator Outlet	$CO_{2\,(dry)}$	93.4	89	96.8	85.5	88.8	85.5	89.9	$\%_{(dry)}$
	O_2	0.38	1.1	3.2	2.3	1.9	2.4	1.1	%
	H_2O	7.2		8.79	10.85	6.56	6.74		%

6.4.4 Review of Pilot Key Performance Parameters

Much was learned during the parametric and continuous operation periods that can be related to the key performance parameters previously presented.

- Reactor pressure drop: Bed levels in the adsorber and regenerator are dependent on the fluidization velocities, the mass of sorbent in the bed, and the temperature of the bed. The total mass of sorbent in the system was increased beyond the design capacity to compensate for the lower velocities and bed levels being used. Evidence of this is shown in the higher than design pressure drop through the adsorber, as measured at the exit of the flue gas blower (BLW-101). The design blower outlet pressure was 5.9 psi, as shown in Table 19, but operations maintained a pressure of 6-7 psi continuously.

- Successful isothermal operation of the adsorber and regenerator: Isothermal operation was approached in the adsorber during pilot operation. The uppermost fluidized bed was warmer at the bottom where sorbent from the regenerator was being added, and it was difficult to maintain the temperature from the bottom to the top of the uppermost bed much closer than within about 5°C of the set point in this dynamic system, as shown in Table 19. In the regenerator, the temperature in the topmost location was slightly lower than the regenerator temperature, but that the total mass of sorbent in the regenerator was close enough to the 120 °C design temperature.
- CO_2 delta loading: The delta CO_2 loading across the adsorber based on CO_2 measurements at the inlet and outlet of the adsorber appeared to be capped at 5.4 g CO_2/100 g sorbent. Results from pilot testing indicate that an overall working capacity of 7 g CO_2/100 g sorbent between the regenerator outlet and adsorber inlet was likely achieved, with the difference between the overall working capacity and the delta loading across the adsorber associated with CO_2 adsorbed from the CO_2-rich conveying gas used to transport sorbent from the regenerator to the adsorber.
- H_2O delta loading: Pilot results from manual measurements show the average moisture uptake to be only 0.49 g H_2O/100 g sorbent, substantially lower than the 1.0 g H_2O/100 g sorbent design point.
- Combined enthalpy of reaction/adsorption of CO_2 and H_2O: As described in previously in this section, the operating conditions during the 90% capture period were not representative of optimal conditions based on pilot testing, and the sorbent-to-CO_2 ratio was nearly four times the design conditions. However, there was excellent agreement between the mass flow observations, heat balance, and fundamental sorbent performance further strengthening conclusions drawn from these data.
- Reaction kinetics of sorbent BN: Results from laboratory tests indicate that initial adsorption occurs within seconds and the sorbent should reach 90% of the equilibrium capacity in less than 12 minutes, as shown in Section 3.6.3 and Table 8. After factoring in the impact of CO_2 adsorbed in the transport line from the regenerator, the working capacity and reaction kinetics observed during pilot tests was consistent with laboratory data. The residence time incorporated into the ADAsorb pilot design should be adequate.
- CO_2 partial pressure in the regenerator: The CO_2 partial pressure in the regenerator is a function of the product purity. The CO_2 product purity from the pilot averaged 89.4% on a dry basis during 90% CO_2 removal from the flue gas. Air in-leakage appeared to affect the CO_2 product purity. Assuming air in-leakage is reduced, the resulting flue gas average concentration can be calculated as 92.4% CO_2, which is considerably closer to the design value of 93.4%.

- Optimize process variables:
  - *Temperature (adsorber and/or regenerator)*: As noted above, the top bed of the adsorber deviated the most from isothermal operation, because of the recycle of sorbent at a higher temperature than the design value of 40°C.
  - *Sorbent Circulation Rate*: Due to changes in material handling characteristics of the sorbent, it was impossible to maintain sorbent flow above about 20,000 lb/hr.
- Identify sorbent bed height required to maintain 90% CO_2 capture: As noted previously, 90% CO_2 capture was constrained because of CO_2 adsorption in the transport line between the regenerator and adsorber and/or non-isothermal conditions in the regenerator.

  Bed levels in the adsorber and regenerator are dependent on the fluidization velocities, the mass of sorbent in the bed, and the temperature of the bed. Bed levels under different operating conditions are included in Table 19. From an operational perspective, the bottom adsorber bed was the most critical level to control in the system. If the bottom bed level dropped below the minimum level, the difference in pressures between the adsorber and regenerator would cause a flow of flue gas into the rich sorbent transport line to the regenerator. When that occurred, a large decrease in the adsorber exit flow corresponded to a high regenerator outlet flow, typically resulting in significant sorbent carryover from the regenerator cyclone to the baghouse. Maintaining bottom bed level became the most critical operator control variable for stable operation. Increasing the sorbent circulation rate between the adsorber and the regenerator tended to cause the bottom bed level to drop, as sorbent would exit the bottom bed faster than it could be replenished.

  The highest continuous sorbent circulation rates tended to correspond to lower than design levels in the middle bed, which encouraged sorbent turbulence and filling of the middle to bottom bed drop tube. Setting the level control valve between the middle bed and bottom bed beyond approximately 50% open resulted in a reversal of flow through the drop tube, preventing sorbent flow between the middle bed and bottom bed of the adsorber. At times the reversal was substantial enough to cause the middle bed to fill beyond maximum level and starve the bottom bed of sorbent. In those situations, only a dramatic decrease in adsorber fluidization flow would break the stagnation and allow commencement of sorbent circulation.

  The top adsorber bed level had a direct correlation to the level of sorbent in the lean sorbent cyclone dip leg (CYC-104). In order to promote flow out of the cyclone dip leg, and increase the sorbent circulation rate, the most effective sorbent levels were maintained at much lower than design (18 ft vs 29 ft design). When the adsorber was operated with a high level in the top bed, sorbent circulation rates decreased with a subsequent increase in the potential for sorbent carryover from the adsorber cyclone (CYC-101) to the baghouse.

- Confirm key fluid bed design parameters:
  - *Bed density:* Bed densities in the adsorber and regenerator are included in Table 19 and are dependent on factors such as fluidization velocities, mass of sorbent in the bed, and the temperature of the bed. During pilot testing, the bed density was higher than design, which corresponds to gas flow that was much lower than the design condition. The density in the regenerator was much lower than the design point, which is likely a result of different handling characteristics of the sorbent when hot compared to the design assumption, which was tested at ambient temperature.
  - *Heat transfer coefficient (cooling in adsorber, heating in regenerator):* The steam flow during parametric testing ranged from 3.2 to 3.8 lb high pressure steam/lb CO_2, compared to a design point of 1.4 lb steam/lb CO_2 for conditions with stable bed conditions. The increased steam requirement during pilot testing is likely a reflection of the lower-than-design regenerator bed density and the lower-than-design mass of sorbent in the regenerator.
  - *Riser lift gas requirement*: The carrier gas in the riser from the regenerator to the adsorber (Cyclone-104) was maintained at nominally design conditions. Due to changes in the handling characteristics of the sorbent when hot, pluggage in the cyclone occurred, especially at sorbent circulation rates near or above 20,000 lb/hr.
  - *Standpipe aeration gas requirement:* Fine tuning of the regenerator outlet CO_2 fluidization valves and the Cyclone-104 nitrogen fluidization valves was performed on 6/12/2015. Data indicates that sorbent flow into the adsorber was periodically limited during operation before the fluidization adjustments due to changes in sorbent handling properties at high temperature that was not factored into the system design.
- Assess the experimentally predicted number of adsorber stages (3) required for 90% CO_2 removal with BN sorbent: Based on the modeling and pilot results, three stages appear preferred for optimal operating conditions with sorbent BN. There are trade-offs between the circulation rate and the number of beds, which may factor into overall economics.

- Confirm the overall feasibility of using solid sorbent BN for CO_2 capture from power plant flue gas in a staged fluidized bed adsorber and fluidized bed regenerator: Because of the limitation on capacity, likely caused by using cold, CO_2-rich gas as the transport gas, the pilot did not achieve the full working capacity of sorbent BN. However, modeling suggests that this limitation could be eliminated by changing how the sorbent is conveyed back into the adsorber.

6.4.5 Energy Balance During 20-Hour, 90% Removal Test

The energy produced from the exothermic reaction of CO_2 and amines in the sorbent should equal the energy released in the endothermic reaction when CO_2 is released from the sorbent. Heat is removed from the ADAsorb system to offset the heat produced through reactions and maintain isothermal conditions in the adsorber. Heat is added to the regenerator to drive the CO_2 off the sorbent and overcome the energy required to break the bonds between the CO_2 and the amines and to drive off adsorbed water.

The energy balance for the ADAsorb System was assessed during the 20-hour period of nearly 90% CO_2 capture starting on June 25, 2015. As described in previously in this section, the operating conditions during the 90% capture period were not representative of optimal conditions based on pilot testing, and the sorbent-to-CO_2 ratio was nearly four times the design conditions. The cooling duty across the adsorber during this period was 0.94 MMBtu/hr. The average energy input to the regenerator for the same period of time, based on the enthalpy difference of steam entering and water exiting the regenerator, was 1.1 MMBtu/hr, which represents a small difference. It is reasonable that the heat duty out of the adsorber would be slightly lower than the heat duty into the regenerator for the following reasons:

1. The fluidization gas into the regenerator was hot, recycled, CO_2. There was a negligible duty of heating for the recycled fluidization gas to the regenerator at 120°C (248°F);
2. The ADAsorb pilot is not truly adiabatic and there will be some heat loss.

The cooling duty of 0.94 MMBtu/hr in the adsorber with only 360 lb/hr of CO_2 removal indicates a very high mass throughput of sorbent (circulation rate) with substantial sensible heat duty to change the temperature of the relatively inert substrate and low heat of reaction duty resulting from CO_2 reacting with amines. This observation is very consistent with a low observed working capacity where the heat of reaction between the sorbent and CO_2 becomes a smaller component of the cooling duty requirements.

The observed energy penalty for regenerating the sorbent during this 20-hour testing period averaged 3,150 Btu/lb_m CO_2 captured. This was calculated based upon the mass and enthalpy of steam entering the regenerator and the mass and enthalpy of the condensate leaving the regenerator. Since the pressure, temperature, and phase of the steam and condensate are known, it is possible to calculate the enthalpy values of the steam and condensate. By using the measured flow rate of the steam and the difference in the calculated enthalpy values, it was possible to determine the heat rate in terms of Btu/hr. Taking this value and dividing by the average CO_2 captured rate of 360 lb/hr across the adsorber yielded the result of 3,150 Btu/lb_m of CO_2 captured.

The expected value based upon laboratory characterization of the sorbent and thermal modeling was 1,530 Btu/lb_m CO_2 captured. This laboratory value was calculated using the following equation:

$$Q = \Delta H_{ad,CO_2} + \Delta H_{ad,H_2O} + C_{p,bed}\Delta T + C_{p,CO_2}\Delta T + C_{p,H_2O}\Delta T \tag{16}$$

The sorbent was characterized using direct scanning calorimetry to determine the heat of reaction with CO_2, the heat of reaction with water, and the heat capacity of the sorbent. These parameters along with the CO_2 and moisture loading viewed at equilibrium in the laboratory were used to calculate an expected thermal requirement of 1,530 Btu/lb_m of CO_2 captured. The variance is due to the average working capacity used to calculate the energy penalty. A capacity of 7 g CO_2 / 100 g sorbent CO_2 by weight was used during the design process of the pilot and in the initial thermal energy requirement calculation. Although the working capacity of the sorbent between the regenerator outlet and adsorber outlet is believed to have been 7 g CO_2 / 100 g sorbent, when including the negative impact of CO_2 adsorption in the transport line between the regenerator and adsorber, the working capacity across the adsorber was limited to 5.4 g CO_2/100 g sorbent, as indicated in Section 6.4.2.

When the sorbent working capacity was calculated using the average CO_2 removal in the adsorber divided by the average sorbent circulation rate for the 20-hour 90% CO_2 capture test, a value of 2.2 g CO_2 / 100 g sorbent was calculated. Using this sorbent working capacity, which was the average value, observed over this particular test duration, the recalculated thermal energy penalty was 3,000 Btu/lb_m, indicating that the thermal properties of the sorbent were well understood and characterized. This also indicated excellent agreement between the mass flow observations, heat balance, and fundamental sorbent performance further strengthening conclusions drawn from these data.

6.4.6 Measurement Uncertainties

Several measurement uncertainties were identified during the project. One of the largest uncertainties was related to the sorbent circulation rate, which was calculated rather than directly measured, and together with the location of the pressure taps used for calculations, left some uncertainty as to whether some sorbent was recirculating back into the regenerator rather than being transported into the adsorber. Another key uncertainty that required modeling to deduce probable behavior was the adsorption of CO_2 in the transport line between the regenerator and the adsorber.

The CO_2 analyzer used to measure the CO_2-rich gas stream required dilution to bring the measurement within range, and did not reliably hold calibrations, which also introduced uncertainty. The average values of CO_2 removed from the adsorber and CO_2 produced from the regenerator were within approximately 20%. This indicated that the data collected was suitable for analysis and for drawing of overall conclusions, but may not have been as precise as desired.

A potential explanation for the difference between the CO_2 removal rate across the adsorber and the CO_2 production rate is that the adsorber used traditional CEMS while the CO_2 product was monitored with a separate emissions monitor with a different calibration range and demonstrated reliability issues associated to calibrations. When the instrument was calibrated according to the manual, it would not accurately read known concentrations of gases. For this reason, a separate calibration procedure was set up to calibrate the instrument and apply a correction factor to the readings each day.

Using the pressure, actual flow rate measurement, temperature, and gas composition collected by the CEMs at the inlet and outlet of the adsorber, a mass flow rate could be calculated. Manual moisture measurements were taken to convert the dry gas CO_2 concentration measurements to wet gas readings for use with the flow meter data. Manual moisture measurements are another source of uncertainty as they can be affected by operators, length of sampling time, and quality of sampling preparation. This is another source of calculation error.

It was not possible to perform enough moisture measurements at a given time to have statistically significant error bars during the test since moisture measurements were taken at the same sampling port as the gas analyzers. This meant that during moisture sampling, the data from the pilot had to be ignored because it did not contain accurate gas compositions. This led to another potential contribution to uncertainty. The adsorber averaged a CO_2 removal rate of 360 lb/hr, while the CO_2 product averaged a production rate of 300 lb/hr. In addition to the uncertainties already described, air in-leakage also likely played a role in contributing to dilution of the CO_2 product on the suction side of the CO_2 blower which would lead to a lower calculated CO_2 mass flow rate. More detailed analysis of the air in-leakage contribution is discussed in Section 6.4.3.

Measurement in the high-concentration CO_2 stream was performed using a Testo 350 fitted with a CO_2 (NDIR) sensor, among other sensors. This sensor is designed to measure 0 to 50 vol% CO_2 with a resolution of 0.01 vol% using infrared measurement principle. Built into the sensor are an absolute pressure measurement, condensate container filling level monitoring and CO_2 absorption filter with filler pack. The Testo 350 has a built-in dilution which can be programmed and plumbed up to act upon any number of sensors. The CO_2 sensor was configured to incorporate a 2 to 1 dilution in order to have the ability to measure up to 100 vol% CO_2.

There are errors associated with the Testo CO_2 cell, the CO_2 calculation and the use of the dilution all within the Testo instrument. The calculated propagated error of the Testo system based on an 89.4% CO_2 product reading is +/-5.75 vol%.

6.4.7 Define and Collect Compression and Sequestration-Specific Information

The characteristics of the CO_2 at the outlet of the capture system, including the pressure and constraints on impurities, are dependent on the intended fate of the CO_2. For example, a low nitrogen content is important for CO_2 used in EOR because nitrogen can decrease oil recovery[41]. In 2012, NETL published "CO_2 Impurity Design Parameters" to provide guidance on a range of impurities in CO_2 for use in EOR or for storage[42]. The NETL-recommended design limit for nitrogen is 1%, if used for EOR and 4% if used for storage directly. Other recommendations for most of the primary gases expected to be present in the CO_2-rich gas from the ADAsorb System are included in Table 23.

The limit for H_2O is heightened when there is a possibility of forming sulfuric acid and of the associated corrosion risk of sulfuric acid. The SO_2 concentration at the inlet of the ADAsorb System is less than 5 $ppm_{v,}$ which will limit the outlet SO_2 concentration and potential for sulfuric acid formation. The corrosion potential is also higher in the presence of free water. Since CO_2 is transported as a supercritical fluid, the solubility limit of CO_2 at 500 bar is 5000 ppm, which is an order of magnitude above the recommended range for water.

Table 23. Recommended Limits for CO_2-Rich Gas Streams[42]

Component	Unit	Range in Literature	
		EOR	Sequestration in Saline Reservoir
CO_2	Vol (%)	90-99.8	90-99.8
H_2O	ppm_{wt}	20-650	20-650
N_2	Vol%	0.01-2	0.01-7
O_2	Vol%	0.01-4	0.001-1.3
SO_2	ppm_v	10-50,000	10-50,000
NOx	ppm_v	20-2,500	20-2,500

The purity of the CO_2-rich gas exiting the pilot system was presented in Table 24. As shown, the CO_2 and O_2 levels were within the recommended range. The elevated levels of water would require additional drying prior to compression and sequestration. If the balance of the gas at the regenerator outlet is nitrogen, there are indications that at some operating conditions, the N_2 content is above the recommended level.

The CO_2 product stream can potentially be diluted or contaminated by at least three mechanisms:

1. Air infiltration or in-leakage on low pressure side of CO_2 blower.
2. Sorption of O_2 and/or N_2 on the CO_2 sorbent.
3. Carryover of flue gas in interstitial spaces of sorbent flowing out of the bottom of the adsorber.

All three issues were investigated to the extent practical after completion of the pilot testing.

1. Leak checking of the pilot was not very practical after testing had concluded and the data were being analyzed. Given the mass balance discrepancy already discussed, it is likely that there was some air leakage. Leakage around the blower driveshaft is possible unless measures are taken to mitigate leakage. No special measures were taken on the pilot blower, so this is likely one source of dilution.

2. If the CO_2 stream dilution were due to only to flue gas carry-over, then the ratio of N_2 and O_2 would be similar in the CO_2 product and in the flue gas. The ratio of N_2 to O_2 in the flue gas was observed to be ~14, while the ratio in the CO_2 product was 8.2. Ambient air has a ratio of N_2 to O_2 of approximately 3.7. It is likely that both air in-leakage and flue gas carryover contributed to dilution of the CO_2 product and limited CO_2 concentration to ~90% on a dry basis.

Models had predicted some interstitial flue gas carryover in the process and determined a ratio of 13 for N_2:O_2 in the CO_2 product stream. The CO_2 product gas stream was predicted to be 94.5% CO_2 with ~5% N_2 on a dry basis. If we assume that the reason for the pilot N_2:O_2 ratio of 8.2, which is about the average of the ratios for air and flue gas, is due to a combination of air and flue gas infiltration, then it is possible to determine their relative contributions. Respectively, these contributions are 55% air and

45% flue gas ingress. Assuming air in-leakage is resolved, the resulting flue gas average concentration can be calculated as 92.4% CO_2, 0.51% O_2, and 7.0% N_2, which is considerably closer to the performance expected.

6.5 Environmental Considerations

The as-produced sorbent selected for use in this project is considered nonhazardous by the OSHA Hazard Communication Standard. Throughout the operation of the pilot samples were extracted from the system and analyzed for RCRA 8 metals to monitor for any accumulation of metals that could lead to reclassification of the used material. The results of this testing, as seen in Table 24, show that no significant accumulation of metals occurred and that the majority of the monitored constituents were below the recordable limit of the testing method.

Table 24. Results of RCRA 8 Testing of Used Sorbent.

Date	Minimum Recordable Limit	10/2/14	4/8/15	5/22/15	5/22/15	5/26/15	5/29/15	6/10/15	6/10/15	Units
Location		ADS	ADS	ADS	REG	ADS	ADS	ADS	REG	
Arsenic	5	-	-	-	-	-	-	-	-	mg/kg
Barium	0.5	2.06	-	-	0.528	-	-	0.046	0.011	mg/kg
Cadmium	1	-	-	-	-	-	-	-	-	mg/kg
Chromium	2.5	-	-	-	-	-	-	0.005	-	mg/kg
Lead	2.5	-	-	-	-	-	-	-	-	mg/kg
Mercury	0.05	-	-	-	-	-	-	-	-	mg/kg
Selenium	2.5	-	-	-	-	-	-	-	-	mg/kg
Silver	2.5	-	-	-	-	-	-	-	-	mg/kg

Total liquid discharge of the system was monitored for total suspended solids, HEM (oil and grease), as well as RCRA metals. The total discharge was the combined discharge of the SO_2 scrubber blowdown, the flue gas cooler condensate, CO_2 cooler condensate, as well as the steam condensate. Results from this testing, seen in Table 25, also indicated that no significant accumulation occurred and that the majority of the monitored constituents were below the recordable limit of the testing method.

Table 25. Results from Environmental Testing of Total Liquid Discharge.

Date	Minimum Recordable Limit	6/10/2015	6/10/2015	Units
Location		D-101	Scrubber	
Arsenic	0.05	-	-	mg/L
Barium	0.005	0.01	0.011	mg/L
Cadmium	0.005	-	-	mg/L
Chromium	0.005	-	-	mg/L
Lead	0.02	-	-	mg/L
Mercury	0.0002	-	-	mg/L
Selenium	0.02	0.02	-	mg/L
Silver	0.005	-	-	mg/L
Zinc	0.01	0.01	-	mg/L
Total S. Solids	4	8.8	9	mg/L
HEM	6	-	-	mg/L

Pilot testing indicates that no special consideration for process waste is required.

6.6 General Observations and Lessons Learned from Pilot Operations

During the parametric testing, it became clear that 90% CO_2 capture at 1 MWe was not achievable with this sorbent in the pilot unit as designed and operated. Unfortunately, the importance of a critical operating condition, using CO_2-rich gas that was cooler than the regenerated sorbent to transport sorbent from the regenerator to the adsorber, was underestimated during pilot testing because of the misinterpretation of laboratory data concerning sorbent kinetics. Without reducing the impact of CO_2 in the sorbent transport line between the regenerator and the adsorber, no variation of any other parameters that were tested led to substantial improvements in performance. A summary of general observations and considerations for improvements during subsequent testing are provided below.

1. The CO_2 removal rate across the adsorber was limited due to adsorption of CO_2 in the conveying gas between the regenerator and the adsorber. It is possible that significant improvements in performance may be achieved through either heating the carrier gas between the regenerator and adsorber, or using a low-CO_2 carrier gas. As noted, the initial design identified CO_2-lean flue gas as the carrier gas. The risk of oxidation and permanent damage to the amines in the sorbent when amines are exposed to oxygen at elevated temperatures was demonstrated during laboratory testing.
    - *Elevating the regenerator above the sorbent return in the adsorber would allow for gravity feed and eliminate the need for carrier gas between the regenerator and adsorber.*

2. The sorbent flow properties changed as regenerator temperature increased. As the regenerator temperature approached the design set point of 120°C (248°F), various sorbent handling characteristics became apparent such as disruption of consistent sorbent circulation rate as indicated by varying levels of sorbent level in the downcomer of the sorbent return to adsorber cyclone and a "slugging" flow of sorbent back into the adsorber from the regenerator.
    - *Modifying the design of the sorbent outlet of the regenerator to incorporate gravity feed and larger conveying lines could mitigate handing issues associated with elevated temperature sorbent.*
3. Cyclones, loop seals, did legs, and downcomers did not always operate smoothly. Sorbent carryover and reverse flow through the cyclones happened on numerous occasions during periods of operational instability. Additional fluidization installed in the downcomer, which returned sorbent from the regenerator, aided in minimizing sorbent circulation challenges. At times, sorbent was also found plugging the gas return line after Cyclone-104, the cyclone responsible for separating regenerated sorbent from conveying CO_2 and directing the sorbent to the adsorber. The presence of sorbent within this line is indicative of sorbent recirculation to the regenerator having bypassed the adsorber. All cyclones were bypassed at some point due to upset conditions.
4. Automatic blower shutoff programming was added to the PLC as well. This proved very valuable as the PLC would detect pressure drop across the sorbent baghouses and automatically shut down the main flue gas blower and/or the CO_2 recycle blower for the regenerator prior to exceeding the deflagration panel burst pressure of 1.0 psig. On two occasions, excessive sorbent carryover through cyclones and into the baghouses resulted in the release of deflagration panels and subsequent release of sorbent. Reinstallation of the panels was costly and time consuming. With the automatic shutdown programming, system instability that could have led to another panel rupture was safely mitigated.
5. Attempts to increase sorbent inventory were unsuccessful due to an inability to maintain a sorbent inventory of greater than approximately 8,500 lb in the system. Occasionally, it was possible to operate close to 9,000 lb, but often additional sorbent would then be carried out of the system to the baghouse. The designed sorbent capacity of the system is 9,900 lb. The inability to increase the sorbent inventory is likely due to the material handling characteristic changes of the heated sorbent at the exit of both vessels, adsorber and regenerator, where heated sorbent is handled. Sorbent inventory was calculated using a dP measurement across each bed to determine density of each bed and then calculating total mass as a function of bed height and vessel diameter.
6. Due to the upper limit threshold of CO_2 removed across the adsorber in terms of lb/hr, it was determined that in order to execute testing at 90% CO_2 removal, adsorber flow would have to be reduced significantly from 1 MWe to 0.2-0.25 MWe.
7. The Testo multigas combustion tuning analyzers measuring CO_2 and O_2 concentration at the adsorber inlet and outlet were replaced with conventional CEMS on loan and maintained by the National Carbon Capture Center. Data from the Testo analyzers was difficult to accurately record in data logs and did not provide long term continuous reliable readings.

General recommendations for scaling sorbent-based CO_2 capture technologies from the laboratory to the pilot-scale include the following.

- Plan for more balance of plant issues on first-of-a-kind pilots than expected. Challenges during this program included
    - Inherent instability of fluidized beds
    - SO_2 scrubber operation and subsequent blower issues
    - Electrical grounding issues
    - Unseasonably cold weather and inadequate freeze protection
    - Supply water contamination
    - Analyzer reliability
    - Insufficient fluidization within transfer lines, sorbent carryover
- Conduct additional lab tests, extrapolate performance, and build in contingency. Unique challenges on this program included:
    - Changes in sorbent behavior at high temperatures
    - Misinterpretation of sorbent kinetics from laboratory data, which resulted in overlooking the potential adsorption of CO_2 when using CO_2-rich gas to transport sorbent from the regenerator to the adsorber.

7. Techno-Economic Analysis

Two high level techno-economic assessments (TEA) were completed: one prior to pilot testing using design conditions and one following pilot testing using data collected during the 20-hour, 90% CO_2 capture test. The objective of the TEAs was to determine the economic viability of the emerging solid sorbent technology by comparing it to an established absorption technology at full-scale commercialization. For the TEAs, the benchmark technology selected for comparison was the Fluor Econamine FG Plus^{SM} (Econamine) process. The core equipment costs for the cases Econamine1 and Econamine2, and Sorbent1 and Sorbent2 were identical, respectively, but the method of calculating indirect costs for the final BEC cost of the CO_2 capture facility differed. The “1” options utilized the DOE method to build up costs and indirects[40]. The “2” options began with the same core costs, but built up the estimate using a more comprehensive estimate of indirect costs, based on recent vendor quotations and construction experience. The methodology provided by DOE/NETL was used to calculate a cost of energy (COE) and a levelized COE (LCOE)[40] for case Sorbent1 (note that Econamine1 was taken directly from the DOE report with inflation taken into account).

A comparison of all case studies is provided in Table 27. Based on the results from pilot testing, and extrapolations using the operating model for conditions where adsorption of CO_2 in the transport line between the regenerator and adsorber was mitigated, the TEA utilizing pilot design parameters is valid. For this case, the sorbent-based CO_2 capture resulted in lower cost of electricity increases (as compared to the base case) than the benchmark Econamine based CO_2 capture.

Preliminary estimates indicate that a PC plant featuring the ADAsorb Process could lead to improved economic viability as well as a greater technical performance than that of the MEA process published in the DOE Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity, Rev. 2.23. The COE is lowered by three percent and the net plant efficiency is increased by 13 percent for the initial study. Compared with the PC plant without carbon capture (Case 9) published in the DOE report, the PC plant featuring ADA's solid sorbent technology results in a COE increase by 80 percent from 67.6 (escalated to 2015 dollars) to 121.8 mills/kWh, which is greater than the DOE target of 35 percent increase in COE.

The following list was generated to highlight the comparison of the key economic considerations.

- Capital costs, fuel costs and CO_2 transport, storage, and monitoring costs of the sorbent-based process were lower than those of the MEA case, respectively. Specifically, to generate 550 MW_{net} with 90% CO_2 capture, the gross power plant was smaller when sorbent-based CO_2 capture was used due to the lower thermal input required to operate the CO_2 capture facility.
- The operating and maintenance costs of the sorbent-based CO_2 capture process were higher than those of the MEA process, mostly as a result of a high sorbent cost and consumption for an initial fill.
- The electric requirement for the compression of CO_2 is significant for both aqueous amine and sorbent-based CO_2 capture. However, because the gross power plant with sorbent-based capture is smaller, less CO_2 must be compressed and the related costs are nominally less.
- COE and LCOE of the sorbent-based process were lower than those of the aqueous MEA system due to a lower capital costs, fuel costs, and CO_2 TS&M costs, offsetting higher O&M costs.
- The cost of electricity and levelized cost of electricity with sorbent-based CO_2 capture were calculated to be 113.3 and 143.6 mills/kWh, respectively.

A complete equipment list including CO_2 capture process equipment, sorbent handling, and CO_2 compression equipment was compiled. Additionally, a process flow diagram was developed and illustrated requirements for integration with an existing coal-fired power plant. These details can be found in the report included in the appendix.

In addition to costs, there are also environmental benefits for the sorbent-based CO_2 capture. For example, the coal supply to the PC plant is linearly related to the power plant gross thermal output. Because the gross thermal output for the power plant with sorbent-based CO_2 capture is ~11 percent less than that for the Econamine process, less coal is burned and thus, less Hg, SO_2, SO_3, NO_x, etc. are created.

Similarly, the raw water consumption of the power plant with the solid based process (~6.5M gallon/D) is 20 percent lower than that of the Econamine process (~8.1M gallon/D). This can be attributed to the fact that the gross power plant is smaller when the sorbent-based CO_2 capture process is implemented. Because water is likely to become an increasingly scarce resource in the future, less water consumption could play an important role in the CO_2 capture technology selection process.

A significant environmental concern is related to the volatile emissions of amines. At the regeneration temperature of 120°C, the vapor pressure of 3.5 M aqueous MEA is approximately 0.0725 psi; emissions of amines could be a significant environmental concern for aqueous CO_2 capture systems. Although the sorbent has been degraded in the laboratory using excessively high temperatures, no measureable volatile emissions were recorded using a mass spectrometer, and no degradation of sorbent exposed to flue gas and temperature cycles was noted from sorbent samples removed from the pilot. It is possible that the amine on the selected sorbent remains covalently attached to the surface of the substrate even when being converted to urea, heat stable salts, or being otherwise degraded.

Solid waste disposal in the sorbent-based process was determined to be non-hazardous as a result of EPA Method 1311 screening which indicated that heavy metals including mercury and selenium were not concentrated in the sorbent. As a result, no special handling or additional costs or equipment was determined to be necessary for sorbent disposal. These results are provided in Section 6.6. Thus, if Sorbent BN were to spill either in transport or at the power plan, unlike an aqueous solution the sorbent will not seep into the soil. The sorbent could more easily be cleaned up without serious environmental consequences. The sorbent is not considered hazardous by the OSHA Hazard Communication Standard, thus waste disposal should be in accordance with the plants locale environmental controls.

Alternative heating and cooling options were also assessed. For example waste heat from compressors and other equipment, use of low quality steam, and potential benefits from alternate sources of heat were evaluated. Incorporating a heat exchanger for the sensible heat of the sorbent and utilizing all heat integration options it was determined that the heat rate could be reduced by 13% from the base design case. In general, the capital costs of these modifications almost always outweighed the benefits with the exception of a cross heat exchanger. Results of individual parameter changes which resulted in decreased heat rate are displayed in Table 26. More detail may be found in the final report for DOE project DE-FE0012914.

Table 26. Heat Rate, COE, and LCOE Changes Resulting from Thermal Optimization

Parameter	Heat Rate	Δ Heat Rate (% from Base Design)	Cost of Electricity ($/MWh)	Δ Cost of Electricity (% from Base Design)	Levelized Cost of Electricity ($/MWh)	Δ Levelized Cost of Electricity (% from Base Design)
Base Design	13364	0.00%	154	0.00%	195	0.00%
FG FWH	13302	-0.46%	170	10.39%	215	10.26%
FG Air	13259	-0.79%	155	0.65%	195	0.00%
Comp FWH	12927	-3.27%	156	1.30%	197	1.03%
Comp REG	12472	-6.67%	150	-2.60%	189	-3.08%
XHTX	12156	-9.04%	148	-3.90%	187	-4.10%
All HI	11625	-13.01%	162	5.19%	205	5.13%
XHTX & Comp REG			147	-4.55%	186	-4.62%

The abbreviations included in Table 26 include the following.

- Base Design: Case without XHTX or heat integration at optimal adsorber and regenerator operating temperatures.
- FG FWH: Heat integration option where heat from the flue gas cooler is used to heat boiler feedwater through the addition of FWH0 upstream of the existing FWH1.
- FG AIR: Heat integration option where heat from the flue gas cooler is used to pre-heat boiler combustion air.

- Comp FWH: Heat integration option where heat from the CO2 compressors is utilized in heating the boiler feedwater through the addition of a feedwater heater (FWH0) upstream of the existing FWH1.
- Comp REG: Heat integration option with heat from the CO2 compressors being utilized in heating the sorbent entering the regenerator.
- XHTX: Addition of a cross heat exchanger with an effectiveness of 75%, operating between the hot and cool sorbent conveying lines to preheat and precool sorbent before it enters the regenerator and adsorber respectively.
- All HI: All five heat integration options were considered simultaneously.

If data from the 20-hr, 90% capture test is utilized for the TEA, solid sorbents are not competitive with the Econamine. The major causes for poor results are due to low working capacity of the sorbent as previously discussed. A low working capacity means that the difference in loading between rich and lean sorbent is reduced, thus the amount of sorbent required for 90% capture of emitted CO_2 grows. This leads to multiple equipment trains for absorption and regeneration, resulting in very high capital costs.

The TEAs for both the design case and the 20-hr, 90% CO_2 removal case are included in the appendix.

Table 27. Performance and Economic Comparison, 2015 Dollars

Description	**Unit**	**Econamine1**		**Econamine2**		**Sorbent1**		**Sorbent2**		**20-hr Pilot Test**	
		BEC as per DOE		BEC based on experience		BEC as per DOE		BEC based on experience		BEC as per DOE	
Gross Power Output	kW_e	672,700		672,700		656,125		656,125		718,738	
Electrical Output – BP Turbine	kW_e	0		0		47,369		47,369		32,525	
Auxiliary Load Summary	kW_e	122,740		122,740		153,534		153,534		201,293	
Net Power Output	kW_e	549,960		549,960		549,960		549,960		549,970	
Thermal Input	kW_{th}	2,102,644		2,102,644		1,865,320		1,865,320		2,383,129	
Coal Flow Rate	lb/hr	614,994		614,995		545,581		545,581		697,032	
Net Plant HHV Efficiency	%	26.2		26.2		29.5		29.5		23.1	
Net Plant HHV Heat Rate	Btu/kWh	13,046		13,046		11,573		11,573		14,785	
CO_2 Capture Profile											
• CO_2 Produced	lb/hr	1,459,838		1,459,838		1,295,067		1,295,067		1,651,831	
• CO_2 Captured	lb/hr	1,313,960		1,313,960		1,165,561		1,165,561		1,486,648	
• CO_2 Emission	lb/hr	145,878		145,878		129,507		129,507		165,183	
Circulating Cooling Water Demand†	gpm	540,000		540,000		399,276		399,276		487,120	
Raw Water Consumption	1000 gal/D	8,081		8,081		6,460		6,460		7,802	
CO_2 Capture Facility											
• Regen. Energy	Btu/lb CO_2	1,530		1,530		1,139		1,139		2,042	
• Total Energy	MMBtu/hr	2,010		2,010		1,328		1,328		3,036	
• Steam Flow	lb/hr	2,219,454		2,219,454		1,429,730		1,429,730		2,709,721	
• Sorbent Amount	lb/hr	N/A		N/A		16,610,220		16,610,220			
Capital Cost											
Bare Erected Cost	$	1,415,831,000		1,931,818,999		1,294,480,737		1,662,613,151		3,377,994,721	
Total Plant Cost	2015$	1,847,389,582		2,613,159,050		1,673,687,989		2,218,697,156		4,724,953,270	
Total Overnight Cost	2015$	2,266,414,272		3,204,167,189		2,071,764,559		2,741,840,062		5,991,041,269	
Total As-Spent Cost	2015$	2,583,712,271		3,652,750,595		2,361,811,597		3,125,697,670		6,355,138,125	
Operating & Maintenance Cost		**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**
Total Fixed Op Costs	$M		61.0		81.2		56.8		71.2		141.3
Total Var. Op. Costs	**$M**	**3.1**	**42.8**	**3.1**	**51.2**	**13.0**	**73.2**	**13.0**	**62.9**		**132.0**
Fuel Costs	**$**	**99,798,307**		**99,798,307**		**87,671,276**		**88,534,219**		**113,110,975**	
CO_2 TS&M Costs	$	**27,112,268**		**27,112,268**		**23,815,777**		**24,050,194**		**30,675,509**	
COE contribution (2015$)	**mills/kWh**	**125.0**		**160.3**		**121.8**		**143.3**		**283.9**	
• Capital Costs	mills/kWh	68.7		97.0		62.8		83.1		181.4	
• Fixed Op. Costs	mills/kWh	14.9		19.9		13.9		17.4		34.5	
• Variable Op. Costs	mills/kWh	10.4		12.5		17.8		15.4		32.9	
• Fuel Costs	mills/kWh	24.4		24.4		21.4		21.6		27.6	
• CO_2 T&SM Costs	mills/kWh	6.7		6.7		5.8		5.9		7.5	
LCOE (2015$)	**mills/kWh**	**158.7**		**203.2**		**154.3**		**181.6**		**359.9**	

8. Market Analysis

8.1 Market Drivers

CO_2 capture or avoidance is only pursued in two cases:

1. Forced emissions regulation currently pending;
2. CO_2 production as a bulk commodity product.

In the post-combustion CO_2 capture market, the predominant market driver will be CO_2 emissions regulations. With the recently concluded COP21 meetings in Paris and proposals including the EPA's Clean Power Plan, regulation of CO_2 seems inevitable; however, it is not yet law in the United States as of December, 2015. For this reason, there currently does not exist a market driver for CO_2 capture as it is far cheaper to emit CO_2 than to capture it.

However, regulation will certainly be finalized at some point in time. Exactly when and how is not yet clear. Various potential market drivers could spur development or drive markets towards other technical solutions depending upon how the regulatory framework of CO_2 emissions is ultimately promulgated.

In the meantime, the only current market driver for CO_2 capture is subsequent sale of the CO_2 as a commodity chemical. The market for CO_2 as a bulk chemical product is found in industries such as enhanced oil recovery (EOR), pulp and paper, welding, food and beverage, and medical applications. According to Ibis World's CO_2 market report, this market provides an average price of $40/ton of CO_2 consumed with an annual domestic market of over $600 million per year not including geologic extraction.

In addition, with a contracting food and beverage industry reducing soda production and the bulk of CO_2 for enhanced oil recovery (EOR) currently provided from geologic reservoirs, this market for CO_2 as a product is likely a shrinking market. This is further exacerbated by a current WTI price of $38/barrel greatly reducing the demand for CO_2 for EOR.

8.2 Applicability of the ADAsorb Technology

The ADAsorb technology is most applicable in post-combustion capture situations where an existing plant would be retrofit for CO_2 capture or for construction of a new power plant where only partial slipstream capture was required. In these applications, the technology may be integrated with the plant to achieve from 0% to 90% CO_2 capture and may be tailored to the needs of any given carbon-fueled power plant, including natural gas-fired units if a sorbent suited for low CO_2 concentrations is selected.

In one example, if partial capture of 30% is required for an existing plant under future regulations such as the Clean Power Plan, the ADAsorb technology with a sorbent that utilizes a 5 g CO_2 / 100 g sorbent working capacity and 12 minute adsorption time and costs $5 per lb, may facilitate compliance for as low as \$25/ton of CO_2 avoided. Depending upon carbon taxes or future regulation, such slipstream partial capture opportunities may have significant applicability if sorbent technology also improves.

Other potential applications include CO_2 separation from natural gas. Natural gas cleanup is the primary energy-related market for CO_2 separation and certainly the largest in terms of tons of CO_2 processed or removed as a contaminant. While not the focus of this particular program, the technology may potentially be of use in other fields as markets develop.

8.3 Market Potential

As discussed, the current market potential is very limited for CO_2 capture from electric generating stations until regulations are in place. In the meantime, the natural gas clean-up market may be examined as a potential use for the CO_2 separation technology.

Data on CO_2 separation and venting is very difficult to obtain and not readily advertised by industry. Separation and venting is also very dependent upon gas composition as some fields do not require CO_2 removal to meet CO_2 pipeline quality specifications of less than 2% by volume, while other fields may produce gas that is to 30% to 40% CO_2 by volume. In order to assess the size of the market, some assumptions had to be made. Historic data from 2005 indicated that approximately 1.6% of the volume of natural gas produced in the U.S. was CO_2 that was vented to the atmosphere. This is a reasonable approximation as the IPCC assumes that 1.5% to 2% of natural gas produced is CO_2 that is vented to the atmosphere.

Given the assumptions that 1.6% of natural gas produced is CO_2 that is eventually vented and using current natural gas production according to IEA, the approximate amount of CO_2 separated and vented is 31 million tons per year.

In order to determine the market size of the current CO_2 separation market for natural gas, economic assumptions had to be made. Adapting data provided by the National Energy Technology Laboratory for post-combustion capture using MEA to estimate the operational costs of separation of CO_2 from flue gas yields a result of \$17.41/ton of CO_2 separated. This is likely a reasonable assumption because MEA separation processes were developed for this application and later adapted to post-combustion CO_2 capture. This assumption does not include the additional capital cost of the equipment. Thus, it is reasonable to assume that the CO_2 separation market for natural gas substantially exceeds \$540 million/year domestically when equipment costs are included.

9. Conclusions

Overall, following start-up and commissioning challenges that are often associated with first-of-a-kind pilots, the pilot plant operated as designed and expected with a few key exceptions. The two primary exceptions were associated with handling characteristics of the hot sorbent that were outside of design specifications, which were collected at ambient temperature and CO_2 adsorption in the transport line between the regenerator and adsorber that pre-loaded the sorbent with CO_2 prior to entering the adsorber. In general, the pilot program was a success and indicates that solid sorbent-based post-combustion capture can be utilized to achieve 90% CO_2 capture from coal-fired power plants. Specific conclusions regarding the ADAsorb pilot system operating with Sorbent BN include:

- The ADAsorb pilot could not be operated at the design sorbent circulation rate with the regenerator at design temperature. The key limitation was pluggage in the sorbent outlet of the regenerator, and the cyclone and conveying lines between the regenerator and adsorber. During circulation tests with the regenerator operating at adsorption temperature, sorbent circulation could be maintained, thus, the operational challenges seem to be caused by changes in sorbent handling characteristics when the sorbent was heated.

- Pilot results in conjunction with laboratory studies of sorbent kinetics and an operational model of the adsorber suggest that the sorbent was adsorbing CO_2 from the CO_2-rich gas used to convey sorbent between the regenerator and adsorber. It is also possible that minor non-isothermal conditions (5°C below design condition) in the regenerator could have contributed to the higher than expected CO_2 loading on the sorbent entering the adsorber. The operating model indicated that design performance could be achieved if regeneration at 120°C could be achieved, pre-loading of CO_2 onto the sorbent could be eliminated and operations weren't constrained by sorbent handling issues.

    - *Pilot performance could be validated through additional pilot tests with a focused plan and minor modifications to the pilot to accommodate sorbent handling issues.*

- Preliminary TEA estimates indicate that a PC plant featuring the ADAsorb Process could lead to 3% reduction in COE and an increased net plant efficiency of 13% compared to the benchmark MEA. Compared with the PC plant without carbon capture (DOE's Case 9), the PC plant with an ADAsorb Process results in a COE increase by 80 percent from 67.6 (escalated to 2015 dollars) to 121.8 mills/kWh, which is greater than the DOE target of 35% increase in COE.

- An additional reduction in the COE of 4.55% could be achieved through heat from the CO_2 compressors being utilized in heating the sorbent entering the regenerator, and the addition of a cross heat exchanger with an effectiveness of 75%, operating between the hot and cool sorbent conveying lines to preheat and precool sorbent before it enters the regenerator and adsorber respectively.

    - *The addition of a cross heat exchanger could also mitigate the potential impact on isothermal operation of the regenerator, as modeled by the CCSI team, and potential desorption of CO_2 in the top bed of the adsorber through localized heating from hot sorbent returning from the regenerator.*

- Laboratory data can be very valuable for predicting performance of larger scale systems when appropriately applied. Interpretation of kinetic rates, mass loading, and sorbent performance including changes in handling characteristics can be well characterized across a wide variety of laboratory conditions. However, it is critical that appropriate metrics be used to scale up laboratory performance to practical system performance.

  - Handling characteristics of sorbents may change under temperature, pressure, loading, and handling conditions and should therefore be examined under bench scale conditions before piloting.
  - Sorbent thermal properties such as heat of reaction, sensible heat capacity, etc. can be effectively measured in a laboratory and then used for design or predictive heating/cooling requirements in a full scale system.
  - Kinetics of adsorption or desorption can be observed in the laboratory, but laboratory conditions must reproduce expected full-scale conditions in order for laboratory kinetic data to be quantitatively used for design and scale-up.

- The solid sorbent tested during this program adsorbed CO_2 selectively from coal-derived flue gas without concentrating other harmful trace elements such as mercury or selenium.
- The market is unclear, but the process has been validated as a basic level. If the market develops, further process improvement and sorbent development is warranted.

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11. Appendix: Techno-Economic Analysis

EVALUATION OF SOLID SORBENTS AS A RETROFIT TECHNOLOGY FOR CO_2 CAPTURE: PRELIMINARY TECHNO-ECONOMIC ASSESSMENT OF SOLID SORBENTS FOR POST-COMBUSTION CO_2 CAPTURE

Topical Report
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Abstract

Capturing and geologically storing the carbon dioxide (CO_2) from point source emitters may be one of the best options for controlling anthropogenic CO_2 emissions. However, currently there are no commercially available options for post-combustion CO_2 capture; in addition, the industry benchmark, aqueous monoethanolamine (MEA) used in a temperature swing process, may be excessively energy intensive and expensive. Therefore, the US Department of Energy National Energy Technology Laboratory is supporting the development of alternative CO_2 capture technologies.

Under cooperative agreement DE-FE0004343 development of a CO_2 capture process based on temperature swing adsorption using solid sorbents is being continued. The goals for the sorbent-based process are to significantly reduce the costs, energy penalty, and environmental impacts of CO_2 capture. One of the first steps of the project was to identify a potential commercial-scale process for CO_2 capture with solids and complete a high level assessment of the techno-economic feasibility; the results of this effort are the focus of this topical report. Based on the commercial-scale process a 1 MW pilot will be designed, built, and operated. Once pilot testing is complete a more in depth techno-economic analysis will be completed based on the pilot-scale test results.

The high level techno-economic assessment was completed. The sorbent based CO_2 capture resulted in lower cost of electricity increases than the benchmark Econamine based CO_2 capture. The following list was generated to highlight the comparison of the key economic considerations:

- Capital costs, fuel costs and CO_2 transport, storage, and monitoring costs of the sorbent-based process were lower than those of the MEA case, respectively. Specifically, to generate 550 MW_{net} with 90% CO_2 capture, the gross power plant was smaller for when sorbent-based CO_2 capture was used due to the lower thermal input required to operate the CO_2 capture facility.
- The operating and maintenance costs of the sorbent-based CO_2 capture were higher than those of the MEA process, mostly as a result of a high sorbent cost and consumption for an initial fill. With the data that will be collected during 1 MW pilot testing it may be possible to reduce the overall mass of sorbent required in the system and, thus, reduce the initial fill.
- The electric requirement for the compression of CO_2 is significant for both aqueous amine and sorbent-based CO_2 capture. However, because the gross power plant with sorbent-based capture is smaller, less CO_2 must be compressed and the related costs are nominally less.
- COE/LCOE of the sorbent-based process were lower than those of the aqueous MEA system due to a lower capital costs, fuel costs, and CO_2 TS&M costs, offsetting higher O&M costs.

- The cost of electricity and levelized cost of electricity with sorbent-based CO_2 capture was calculated to be 113.3 and 143.6 mills/kWh, respectively.

In addition to costs there are also environmental benefits for the sorbent-based CO_2 capture. For example, the coal supply to the PC plant is linearly related to the power plant gross thermal output. Because the gross thermal output for the power plant with sorbent-based CO_2 capture is ~11 percent less than that of the Econamine process less coal is burned and thus, less Hg, SO_2, SO_3, NO_x, etc. are created.

Similarly, the raw water consumption of the power plant with the solid based process (~6.5M gallon/D) is 20 percent lower than that of the Econamine process (~8.1M gallon/D). This can be attributed to the fact that the gross power plant is smaller when the sorbent-based CO_2 capture process is implemented. Because water is likely to become an increasingly scarce resource in the future, less water consumption could play an important role in the CO_2 capture technology selection process.

Another significant environmental concern is related to the volatile emissions of amines. At the expected regeneration temperature the vapor pressure of 3.5 M aqueous MEA is approximately 0.0725 psi; emissions of amines could be a significant environmental concern for aqueous CO_2 capture systems. Although the sorbent has been degraded in the laboratory using excessively high temperatures, no measureable volatile emissions were recorded using a mass spectrometer. It is possible that the amine on the selected sorbent remains covalently attached to the surface of the substrate even when being converted to urea, heat stable salts, or being otherwise degraded.

Another environmental consideration is related to potential spills. If the proposed sorbent were to spill either in transport or at the power plan, unlike an aqueous solution the sorbent will not seep into the soil. The sorbent could more easily be cleaned up without serious environmental consequences. The sorbent is not considered hazardous by the OSHA Hazard Communication Standard, thus waste disposal should be in accordance with the plants locale environmental controls.

The sorbent-based CO_2 capture process resulted in a lower energy penalty and lower overall increase in the cost of electricity. However, the improvements versus aqueous MEA are not enough to meet the DOE's cost and COE goals. With further improvements it is possible that the sorbent-based CO_2 capture process may be able to make significant progress towards the DOE goals. Potential areas of improvement include:

- Heat integration
    - Heat recovery between the rich and lean sorbent – this must be accomplished in such a way that the energy recovery is significant enough to offset the added capital costs for the heat exchangers
    - Integration within the power plant – using heat from the power plant or the CO_2 compression system to reduce the overall energy penalty
- Reduce capital costs
- Improve sorbent performance and characteristics
    - Working CO_2 capacity
    - Attrition resistance
    - Cost
    - Thermal conductivity
    - Moisture loading
    - Etc.
- Optimize steam condition at IP/LP crossover to eliminate a need of the BP turbine and to integrate into the steam cycle in a more efficient manner.

A unique process for CO_2 capture using solid sorbents has been developed. In the absence of large scale (i.e. pilot-scale) testing to date, it was necessary to use laboratory results or make assumptions regarding the equipment and sorbent performance under the expected operating conditions. To fully capture the true range of operation and fully assess synergistic effects of the entire system it is important to continue this work to the 1 MW pilot scale to validate the following key performance parameters:

- CO_2 delta loading
- H_2O delta loading
- Enthalpy of reaction/adsorption of CO_2
- Enthalpy of reaction/adsorption of H_2O
- Reaction kinetics
    - CO_2 uptake limited by kinetics in the adsorber
    - CO_2 release limited by mass diffusion in the regenerator
- Effective heat transfer coefficients
    - Cooling water requirements
    - Steam requirements
- CO_2 partial pressure in the regenerator
- Reactor pressure drop
- Physical attrition rate
- Successful isothermal operation
- Optimize process variables:
    - Temperature
    - Sorbent Circulation Rate
- Identify sorbent bed height required to maintain 90% CO_2 capture

Using results from the 1 MW pilot test a more in depth techno-economic assessment can be completed. If the process will not meet the DOE's goals related to the cost of electricity increase as described in this report, sorbent properties and process conditions that would result in an acceptable cost increase will be identified.

Table of Contents

1.0 Executive Summary

Under cooperative agreement DE-FE0004343 development of a CO_2 capture process based on temperature swing adsorption using solid sorbents is being continued. A preliminary techno-economic assessment was completed based on the developing sorbent-based CO_2 capture process. From an economic standpoint the sorbent-based CO_2 capture process outperformed the benchmark aqueous monoethanolamine (MEA) process. Note that the Fluor Econamine FG Plus^{SM} was used as the benchmark MEA-based process. Specifically, the capital costs, fuel costs and CO_2 transport, storage, and monitoring costs of the sorbent-based process were lower than those of the aqueous amine case. This can be attributed to the fact that to generate 550 MW_{net} with 90% CO_2 capture, the gross power plant was smaller for when sorbent-based CO_2 capture was used due to the lower thermal input required to operate the CO_2 capture facility.

In addition to costs there are also environmental benefits demonstrated by the sorbent-based CO_2 capture. Because the gross thermal output for the power plant with sorbent-based CO_2 capture is 11 percent less than that of the Econamine process less coal is burned and thus, less Hg, SO_2, SO_3, NO_x, etc. will be created. Similarly, the raw water consumption of the solid based process (~6.5M gallon/D) is 20 percent lower than that of the Econamine process (~8.1M gallon/D). Another potential environmental concern is related to the volatile emissions of amines. At the expected regeneration temperatures the vapor pressure of 3.5 M aqueous MEA is approximately 0.0725 psi. Although the sorbent has been degraded in the laboratory using excessively high temperatures, no measureable volatile emissions were recorded using a mass spectrometer.

Although the sorbent-based CO_2 capture process outperformed aqueous MEA in several key areas, the costs are projected to be higher than the goals set forth by the US Department of Energy (DOE). Therefore, several potential improvements have been identified. The potential feasibility for these improvements will be assessed by operation of a 1 MW pilot and a follow up techno-economic analysis based on the pilot results.

2.0 Introduction

Capturing and geologically storing the carbon dioxide (CO_2) from point source emitters may be one of the best options for controlling anthropogenic CO_2 emissions. Aqueous monoethanolamine (MEA) is generally accepted as benchmark to which developing CO_2 capture technologies can be compared. While it has not been demonstrated on pulverized coal (PC) fired flue gas at the commercial scale, aqueous MEA has been used extensively for CO_2 removal from natural gas and flue gas with the intention of using it for food applications, enhanced oil recovery, etc. Although commercial-scale CO_2 capture from coal-fired power plant flue gas with MEA is likely to be technically feasible, it will also likely be expensive and energy intensive.[1] The DOE, specifically the National Energy Technology Laboratory (NETL), is supporting research and development efforts for CO_2 capture technologies that may offer technical and/or economic advantages when compared to aqueous MEA.

Under cooperative agreement DE-FE0004343 development of a CO_2 capture process based on temperature swing adsorption using solid sorbents is being continued. The goals for the sorbent-based process are to significantly reduce the costs, energy penalty, and environmental impacts of CO_2 capture. One of the first steps of the development project was to identify a potential commercial-scale process for CO_2 capture with solids and complete a high level assessment of the techno-economic feasibility; the results of this effort are the focus of this topical report. Based on the commercial-scale process a 1 megawatt (MW) pilot will be designed, built, and operated. Once pilot testing is complete a more in depth techno-economic analysis will be completed based on the pilot-scale test results.

This report includes both economic and technical considerations related to two potential CO_2 capture processes. For the purposes of this report the Fluor Econamine FG Plus^{SM} (Econamine) process will serve as a benchmark to which the sorbent-based process is compared. The CO_2 capture processes were compared using the assumption that they were implemented at a 550 MW_{net} coal-fired power plant. An engineering firm was subcontracted to complete the economic assessment, during which it was ensured that all assumptions related to economics were identical for both processes. While the sorbent based capture process is still in the early stages of development there are potential economic and technical advantages, which are identified and explored in this report.

3.0 Methodology

3.1 PLANT AND PROCESS DESCRIPTIONS

A potential CO_2 capture process using solid sorbents was developed. Detailed engineering and an independent techno-economic assessment was completed by an external engineering firm Ultimately, the widespread implementation of any capture technology will hinge on favorable operability and cost competitiveness. To determine whether the sorbent-based CO_2 capture process could offer economic benefits compared to the benchmark aqueous amine systems, the engineering firm completed a high level economic assessment and comparison of the two processes. Two different cost estimates were used to determine the cost of electricity increase associated with these CO_2 capture technologies. One cost approach was provided by the DOE,[1] while the other estimate was based upon a database of cost information received by the engineering firm in evaluating multiple CO_2 capture technologies for other retrofit projects.

In addition to costs, the operability, energy usage, and other performance factors were evaluated for the sorbent based CO_2 capture process. Potential improvements in the sorbent-based CO_2 capture process have been identified. The following sections describe the base power plant used for the analysis, the two CO_2 capture technologies, and the cost calculation methodologies utilized the engineering firm.

3.1.1 Power Plant Description

The techno-economic assessment was applied to the sorbent based CO_2 capture process as if it was implemented at the same power plant as the Econamine process in the cited DOE report.[1] The PC power plant was equipped with the best available control technology (BACT) to comply with environmental targets as summarized in Table 2-1. The flue gas desulfurization (FGD) system used for sulfur dioxide (SO_2) removal was a wet limestone forced oxidative absorber, producing gypsum as a byproduct. A fabric filter was used to control particulate matter (PM). The nitrogen oxides (NO_x) concentration was reduced using a combination of low NO_x burners (LNBs) and over-fire air (OFA) systems in the boiler as well as installing selective catalytic reduction (SCR) upstream of the air heater.

Table 2-1 Case Descriptions

Case	SO_2 Removal	PM Control	NOx Control	CO_2 Capture	CO_2 Sequestration
Econamine	Wet FGD/Gypsum	Fabric Filter	LNBs with OFA and SCR	Amine Absorber	Off-site
Sorbent	Wet FGD/Gypsum	Fabric Filter	LNB with OFA and SCR	Solid Sorbent Fluidized Bed Adsorber	Off-site

Ambient conditions and characteristics of the comparison basis PC plant located for the generic plant in the Midwestern U.S. are listed in Table 2-2. These values were obtained from the DOE/NETL report.[1] Please refer to the complete report for a full description of the power plant.

Table 2-2 Site Ambient Conditions and Characteristics

Parameters	Description
Location	Greenfield, Midwestern U.S.
Elevation (ft)	0
Barometric Pressure (psia)	14.7
Design Ambient Temperature, Dry Bulb (°F)	59
Design Ambient Temperature, Wet Bulb (°F)	51.5
Design Ambient Relative Humidity, %	60
Topography	Level
Size (acres)	300 (30 acres for the PC plant)
Transportation	Rail
Steam Cycle – Single Reheat (psig/°F/°F)	2,400/1,050/1,050
Oxidant	Air
Coal Type	Illinois No. 6
High Heating Value (Btu/lb)	11,666 (as received)
Ash/Slag Disposal	Off-site
Water (H_2O)	Municipal (50%)/Groundwater (50%)
Access	Land locked (having access by rail and highway)
CO_2 Storage	Compressed to 2,215 psia, transported 50 miles and sequestered in a saline formation at a depth of 4,055 feet

One potential advantage of solid sorbents is a reduction in the input of energy required for regeneration, which is discussed in detail in 3.3.1. Since both cases were evaluated on a common 550 MW_{net} output basis, the gross steam turbine output was varied to account for different auxiliary and heat power demand for the two cases. Thermal input generated by burning coal was kept sufficient to supply the steam and electricity demand for both PC plant and CO_2 capture facility (including the energy requirement for solvent/sorbent regeneration).

3.1.2 Aqueous Amine Benchmark CO_2 Capture Process (Econamine)

The Econamine process utilizes aqueous MEA with a proprietary corrosion inhibitor. The solvent captures and releases CO_2 via a temperature swing. To minimize the formation of heat stable salts (HSS) in the solvent, a SO_2 polishing unit using 20% (by weight) sodium hydroxide (NaOH) reduces the flue gas SO_2 concentration to 10 ppmv or less. The flue gas exiting the SO_2 polishing scrubber is pressurized by a fan to overcome a 2 psi pressure drop across the absorber.

The flue gas, with a temperature of approximately 135°F, enters the CO_2 capture equipment downstream of the SO_2 polishing unit and travels upward so that it contacts the MEA solvent counter-currently. At the bottom of the absorber, the rich MEA-based solution is collected and pumped to the regenerator. The CO_2 lean flue gas exits the top of the adsorber and is directed to the power plant stack from the top of the absorber. The CO_2 rich solution is preheated by the CO_2 lean solution from the solvent stripper in the lean/rich cross heat exchanger which significantly reduces the sensible heat requirements of the solvent-based CO_2 capture process. Downstream of the rich/lean heat exchanger the cooled CO_2 lean solution is directed to the absorber, whereas the heated rich solution is directed to the solvent stripper (i.e. regenerator) for CO_2 stripping/solvent regeneration using steam from the intermediate pressure/low pressure (IP/LP) crossover pipe of steam turbine. For sequestration purposes, the CO_2 product gas is compressed to 2,215 psia by a six-stage compressor and dehydrated by using triethylene glycol to lower the CO_2 temperature to 40°F.

There are several concerns related to the use of the aqueous MEA for post-combustion CO_2 capture (note that the list below is not considered to be comprehensive):

- Economic
    - Capital costs
    - Operating costs
- Technical
    - Energy Penalty
        - Sensible heat – partially addressed by implementation of the lean/rich heat exchanger
        - Latent heat of reaction
        - Latent heat of vaporization
    - Environmental Considerations
        - Water usage
        - Corrosivity
        - Fouling, poisoning, and oxidation
        - Volatile emissions

For these reasons it is important to develop alternative options for CO_2 capture. With continued development it is possible that the sorbent-based CO_2 capture process being developed could address several of the concerns listed.

3.1.3 Solid Sorbent CO_2 Capture Process

The solid sorbent-based CO_2 capture process employs the use of a dry sorbent characterized by amine functionalization. Under the current project (DE-FE0004343), this technology will be tested at the 1 MW scale. However, because those tests have yet to be completed the high level preliminary techno-economic assessment was conducted using several assumptions based on public literature and laboratory testing, including:

- Particles physically resemble polystyrene beads with respect to size
- Particle density: 36.6 lb/ft^3
- Heat of reaction is 587 Btu/lb CO_2 adsorbed
- Adsorption temperature is 104°F
- Regeneration temperature is 248°F
- Flue gas temperature after the SO_2 polishing unit is 135°F

The sorbent based process integrated into the PC plant is shown in Figure 2-1. Similar to the Econamine process, the sorbent-based CO_2 capture process is implemented immediately upstream of the power plant stack.

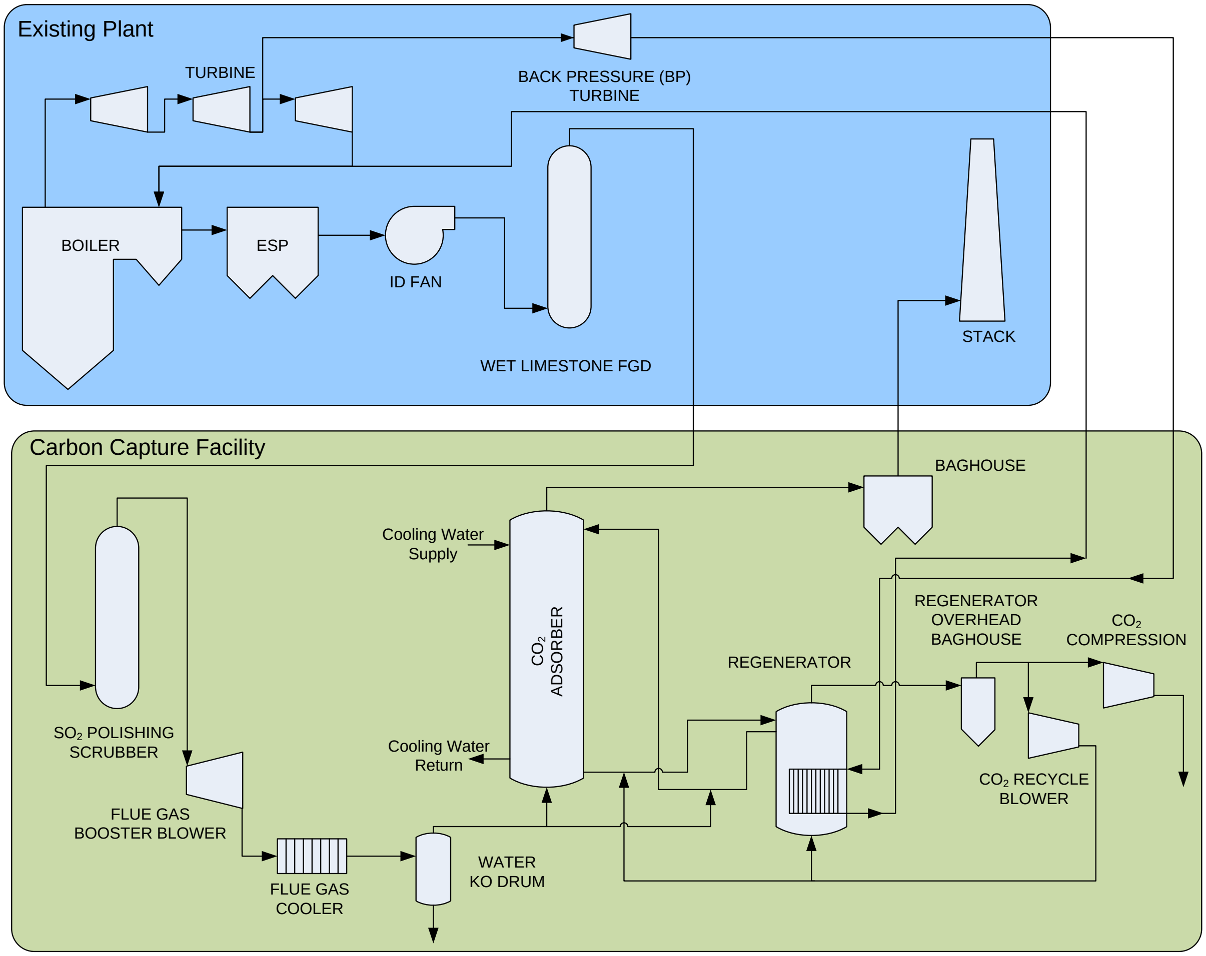

Figure 2-1 Process Flow Diagram of the Subcritical PC Power Plant Retrofitted with a Solid–Based CO_2 Capture Process

The sorbent-based CO_2 capture equipment configuration is provided in Figure 2-2. Flue gas is routed from the existing plant through ductwork to the CO_2 capture facility. The amine functionalization on the sorbent can react to form HSS similar to aqueous amines. Therefore, the SO_2 concentration in the flue gas before entering the CO_2 capture facility must be reduced to levels comparable to those observed in the Econamine process via a similar SO_2 polishing unit. It should be noted an alternative process design which does not require the use of a SO_2 polishing scrubber is also being evaluated. A flue gas blower is necessary to provide sufficient pressure to pass the flue gas through the CO_2 capture equipment, and return it via ductwork back to the stack. In the pilot unit, the flue gas blower increases the pressure of the flue gas by approximately 7 psi, which also causes the flue gas temperature to rise from 135°F to 219°F. It should be noted, the pilot plant is overdesigned to accommodate a series of tests which will be used to quantify the amount of sorbent needed to remove 90% CO_2. Only after the completion of testing, will the full-scale pressure drop of the adsorber be accurately estimated. The flue gas is then directed through a cooler, where the temperature is decreased to 104°F. Water vapor is separated from the flue gas in a knockout (KO) drum.

After the KO drum, the cooled flue gas enters a three-stage fluidized bed adsorber. Each stage in the adsorber is equipped with a bundle of serpentine cooling tubes to maintain isothermal operation. The top bed of the adsorber is responsible for cooling the sorbent from the regeneration temperature (248°F) to the adsorption temperature (104°F), removing the heat generated from the exothermic reaction between CO_2 and the sorbent, and removing the exothermic heat due to the adsorption of the remaining moisture in the flue gas. The cooling coils in the bottom two fluidized beds are responsible only for removing the heat of reaction between CO_2 and the sorbent and the heat adsorption of moisture. Due to this added sensible heat requirement, the top bed depth is larger than the bottom two. The stages are separated by horizontal steel sections, which allow the flue gas entering at the bottom to flow through the adsorber and fluidize the sorbent. The treated flue gas is released at the adsorber top and routed to a series of cyclones followed by a baghouse, which collects any entrained sorbent. Pilot testing will be used to better understand the full-scale requirements for particulate removal downstream of the adsorber and regenerator modules.

The staged fluidized bed was selected for several key performance characteristics, including:

- Maximizing the driving force for mass transfer (i.e. CO_2 uptake) by approaching counter-current gas/solids contacting
- Isothermal operation
- Operation of fluidized bed reactors has been successfully demonstrated in other industries

Standpipes function as passages for sorbent to cascade down from one fluidized bed to the next (top to the bottom), while adsorbing CO_2 from the flue gas flowing upward. As the CO_2 rich sorbent is withdrawn at the bottom, lean sorbent is simultaneously added at the top of the adsorber. The amount of the rich sorbent discharged is controlled by a sorbent slide valve. The CO_2 rich sorbent is pneumatically conveyed with CO_2 to the regenerator via a rich sorbent riser.

In an isothermal (248°F) single-stage fluidized bed regenerator, the rich sorbent releases CO_2 by means indirect steam in an internal heat exchanger. Since the temperature of steam (306°F) extracted from IP/LP crossover section of the PC plant steam turbine is hotter than the regeneration temperature of the solid sorbent (248°F), the expansion of the steam through a back pressure (BP) turbine before entering the exchanger is required to prevent the thermal degradation of the amines impregnated in the solid sorbent and also to generate additional electricity. Saturated steam at 45 psia is obtained at the BP turbine exit where the temperature and pressure of the steam will adjusted accordingly to ensure the amine functional groups are not damaged during the regeneration process.

The higher temperature in the regenerator reverses the reaction between the amine functional groups and the CO_2, thus releasing the CO_2 into the gas phase. The released CO_2 exits the top of the regenerator and is directed through a baghouse. A portion of the CO_2 is routed to the

CO_2 recycle blower, while the remaining CO_2 is sent to the compression plant similar to the Econamine process. The CO_2 recycle blower provides enough pressure for the CO_2 to promote adequate fluidization inside the regenerator and to convey sorbent from the adsorber discharge to the regenerator in the rich sorbent riser. To complete the cycle, the CO_2 lean sorbent from the regenerator is transferred via a sorbent riser, using the conveying media of CO_2 lean flue gas. Four adsorption/regeneration trains are required to process all the flue gas being exhausted from a 550 MW_{net} subcritical coal fired power plant.

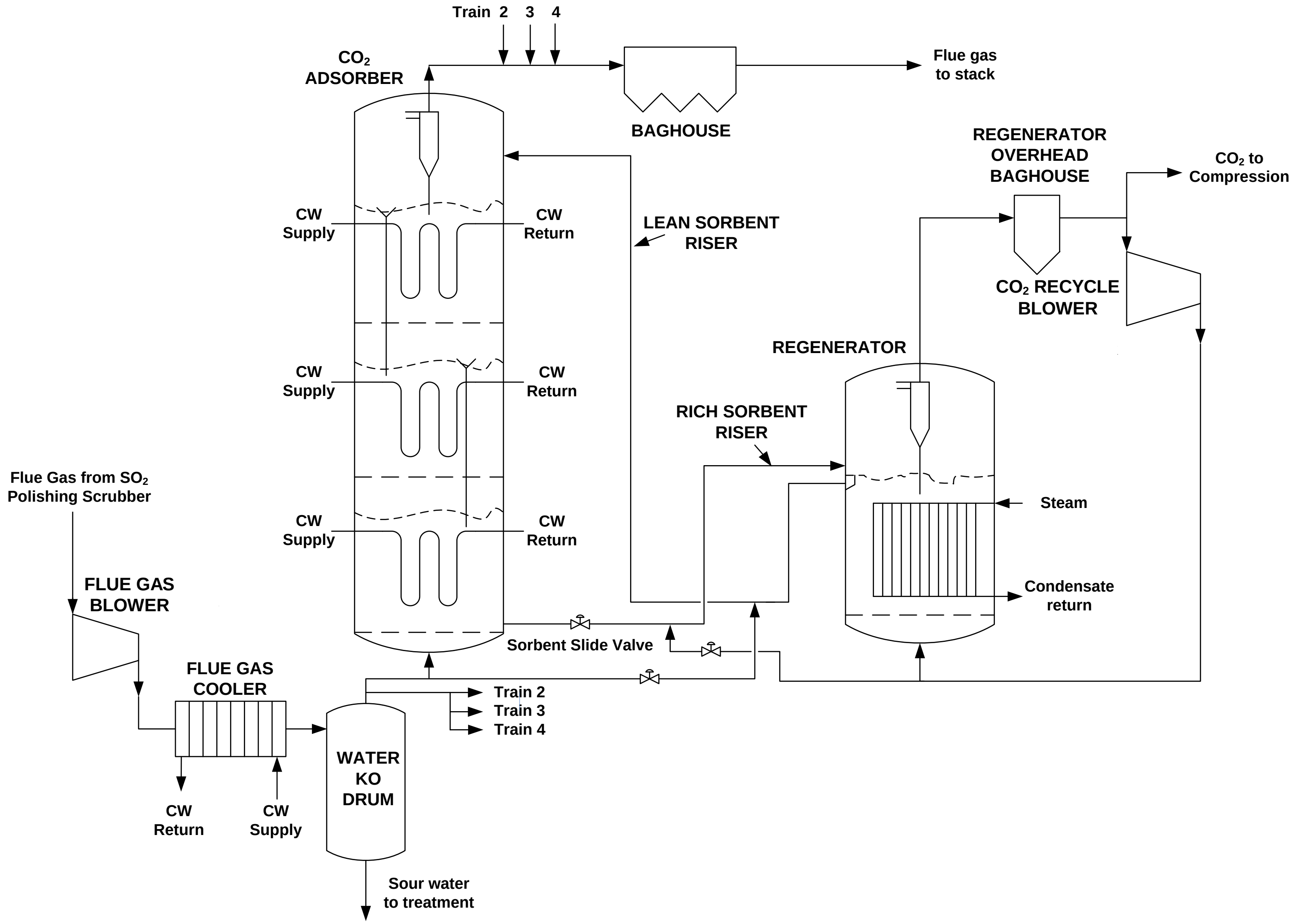

Figure 2-2 Process Flow Diagram of the Solid–Based CO_2 Capture Process

3.2 ECONOMIC ASSESSMENT METHODOLOGY AND ASSUMPTIONS

The Econamine cost and operating information was provided in a report issued by the DOE NETL.[1] In that report the baseline coal-fired power plant with no CO_2 capture is referred to as Case 9 and the same power plant with CO_2 capture using the integrated Econamine process is referred to as Case 10. To ensure a fair comparison between the Econamine process and the sorbent based process, the configuration/conditions (except flow rate) of peripheral units outside the CO_2 capture battery limits were identical, unless some modifications were necessary to meet the requirements of the sorbent-based process. For both cases the power plant generated the same amount of net electricity, 550 MW. Because the energy requirements for the two CO_2 capture processes are different the thermal input to the overall power plant was varied until the net power output was 550 MW.

For each CO_2 capture process two different approaches were used to develop cost estimates. The first approach was based on the methodology described in a report issued by the DOE, entitled “Cost and Performance Baseline for Fossil Energy Plants – Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 2”.[1] The cost estimates based on the DOE methodology are referred to as Econamine1 and Sorbent1. To develop the second cost assessment the engineering firm used vendor quotes and history of past projects from their extensive database. The cost estimates based on engineering quotes and experience are referred to as Econamine2 and Sorbent2.

3.2.1 Capital Costs

As determined by the Association for the Advancement of Cost Engineering International (AACE), the level of cost accuracy for this conceptual design is defined by a Class IV/V estimate, which is approximately in a range of ±50 percent. In order to increase the accuracy of the capital cost estimate, pilot testing should be completed in order to validate numerous process assumptions made in the economic assessment. Pilot data will be used to accurately size the full-scale vessels which will have a proportional impact on the associated capital costs component of the economic assessment. The four levels of capital cost and the estimate basis used in this analysis are given in Table 3-1. Each “level” flows through to the next. The bare erected cost (BEC) is the cost of process equipment plus other costs (e.g. cost of supporting facilities and direct/indirect labor costs related to construction and installation).

Table 3-1 Capital Cost Structure

Level	Cost	Estimate Basis
Bare Erected Cost (BEC) – Overnight cost expressed in base year (2011) dollars	- Process Equipment	
	- Support Facilities (e.g. Offices, Labs)	
	- Direct and Indirect Labor for Construction and/or Installation	
Total Plant Cost (TPC) – Overnight cost expressed in base year (2011) dollars	- BEC	
	- Engineering, Procurement and Construction (EPC) Contractor Services	~ 9.4% of BEC
	- Process Contingency	~ 20% of CO_2 Capture Facility Capital Cost
	- Project Contingency	~ 15% of the Sum of BEC, EPC and Process Contingency
Total Overnight Cost (TOC) – Overnight cost expressed in base year (2011) dollars	- TPC	
	- Preproduction	~ 3% of TPC
	- Inventory Capital	~ 2% of TPC
	- Initial Cost	
	- Land	
	- Owner's Costs	15% of TPC
	- Financing Costs	2.7% of TPC
Total As-Spent Cost (TASC) – Expressed in mixed-year current dollars over the capital expenditure period	- TOC	
	- Escalation/Interest During Capital Expenditure Period	1.140 Multiplier for High Risk Investor-Owned Utility (IOU) Over Five-Year Capital Expenditure Period

Four cost assessments were completed and are described in this report; the cases are listed in Table 3-2. The costs for Econamine1 were taken directly from the DOE report except that they were scaled from 2007 to 2011 U.S. dollars. Based on the DOE report, the power plants for where the Econamine and sorbent-based CO_2 capture processes were both 550 MW_{net}. Because the energy penalty for the two processes is not the same, the gross power plants were different in size. Therefore, the power plant costs were calculated for cases Sorbent1 and Sorbent2 (the power plant costs for Econamine1 and Econamine2 were provided in the DOE report.[1]) For the sorbent cases the capital costs of the PC plant and CO_2 compression unit were obtained by linearly scaling the costs of the Econamine process (escalated from 2007 to 2011

US dollars) proportional to coal usage (i.e. gross power plant size). Total inflation from June 2007 to February 2011 was assumed to be 6.22%.

Table 3-2 Cost Methodology

Case	Cost Estimation Method for BEC of Capture Facility
Econamine1	DOE Methodology (Indirects as per DOE)
Econamine2	Vendor Quotes/Actual Pricing Information (Indirects Based on Experience)
Sorbent1	DOE Methodology (Indirects as per DOE)
Sorbent2	Vendor Quotes/Actual Pricing Information (Indirects Based on Experience)

The cost of the process equipment for the sorbent-based CO_2 capture process, including the SO_2 polishing unit and the BP turbine, were estimated using a combination of vendor quotes and actual pricing information from the engineering firm's internal database. Items specifically excluded from this capital cost estimate are the same as those in the Econamine1 and Econamine2 estimates: taxes, site specific considerations, overtime, and extra premiums due to an EPC contract.

The core equipment costs for Econamine1 and Econamine2 and Sorbent1 and Sorbent2 are identical, respectively, but the method of calculating indirects for the final BEC cost of the CO_2 capture facility differs. The different cost calculation cases and their respective methodologies are listed in Table 3-2. The "1" options utilize the DOE method to build up costs and indirects[1]. The "2" options begin with the same core costs, but build up the estimate using a more comprehensive estimate of indirects, based on recent vendor quotations and construction experience.

3.2.2 Operating and Maintenance Cost Estimates

The operating and maintenance (O&M) costs for cases Sorbent1 and Sorbent2 were developed based on the following constraints and assumptions:

Fixed O&M Costs

- Annual operating labor cost/administrative and support labor were equal to the escalated costs of Econamine1 as provided by the DOE[1]
- Maintenance labor cost/property taxes and insurance were obtained by linearly scaling the escalated costs of Econamine1 proportional to the total plant cost (TPC)

Variable O&M Costs

- An 85% capacity factor was assumed
- Water consumption was estimated using the Econamine1 calculation method.[1]
- Maintenance material cost/annual cost of makeup (MU) water and water treatment (WT) chemicals are obtained by linearly scaling the escalated costs of Econamine1 proportional to the TPC.
- Annual costs of limestone/ammonia/SCR catalyst/fly ash and bottom ash disposal were obtained by linearly scaling the escalated costs of Econamine1 proportional to the gross power plant output
- Initial and annual costs of sodium hydroxide (NaOH)/sulfuric acid (H_2SO_4) were obtained by linearly scaling the escalated costs of Econamine1 A proportional to the power plant gross output. However, it was assumed that the SO_2 concentration must be less than 1 ppm for the solid sorbent-based process. Therefore, the annual NaOH cost was doubled to reflect the greater consumption.
- Consumption of NaOH/Lime/H_2SO_4 for additional waste water treatment was based on the cooling water demand of the CO_2 capture facility.
- A sorbent attrition rate of 0.0025% per cycle (attrition) was assumed. No loss from leakage was taken into account.
- The cost of the sorbent was estimated at US $5/lb.
- Potential revenue from selling CO_2 product was excluded.

3.2.3 CO_2 Transport, Storage, and Monitoring Costs

Supercritical CO_2 from the sorbent based capture facility is transported via pipeline and injected into a deep saline formation for sequestration. Based on the assumptions used to model the transport, storage and monitoring (TS&M) cost for Econamine1, the CO_2 TS&M costs for cases Sorbent1 and Sorbent2 linearly depend on the CO_2 removal rate. The cost for TS&M was calculated to be approximately US $22 M for both cases.

3.2.4 Cost of Electricity

The methodology provided by DOE/NETL was used to calculate a cost of energy (COE) and a levelized COE (LCOE)[1] for case Sorbent1 (note that Econamine1 was taken directly from the DOE report with inflation taken into account). Both COE and LCOE are the revenue received by the generator per net megawatt-hour (MWh) during the PC plant's first year of operation. The difference between the COE and LCOE is that the COE escalates thereafter at a nominal annual rate equal to the general inflation rate of three percent, whereas the LCOE escalates thereafter at a zero nominal annual inflation rate. The calculated COE and LCOE can be used to provide an internal rate of return on equity (IRROE) when the calculation is conducted over the entire economic analysis period using the assumptions and conditions listed Table 3-3.

Table 3-3 Global Economic Assumption

Parameter	Value
Taxes	
Income Tax Rate	38% (Effective 34% Federal, 4% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 years
Contracting and Financing Terms	
Contracting Strategy	Engineering, Procurement, Construction and Management (EPCM)
Type of Debt Financing	Non-Recourse
Repayment Term of Debt	15 years
Grace Period on Debt Repayment	0 years
Debt Reserve Fund	None
Analysis Time Periods	
Capital Expenditure Period	5 years
Operation Period	30 years
Economic Analysis Period	35 years
Treatment of Capital Costs	
Capital Cost Escalation During Capital Expenditure Period (nominal annual rate)	3.6%
Distribution of TOC Over Capital Expenditure Period (before escalation)	5-year period: 10%, 30%, 25%, 20%, 15%
Working Capital	Zero for all parameters
% of TOC that is Depreciated	100%
Escalation of Operating Revenues and Costs	
Escalation of COE (revenue), O&M Costs and Fuel Costs (nominal annual rate)	3.0%

Several assumptions were made to complete the cost assessment. Details on finance structures of the PC plant with CO_2 capture, assuming a high risk investor-owned utility (IOU), are provided in Table 3-4.

Table 3-4 Financial Structure for High Risk IOU

Parameters	%	Current (Nominal) Dollar Cost	Weighted Current (Nominal) Cost	After Tax Weighted Cost of Capital
Debt	45	5.5%	2.475%	
Equity	55	12%	6.6%	
Total	**100**		**9.075%**	**8.13%**

The COE was calculated using equation 1, which was published in the DOE report.[1] The various costs used in this equation were expressed in base-year dollars. The base year, defined as the first year of capital expenditure, for the purposes of the cost calculations the base year was assumed to be 2011. Since the COE, O&M costs, and fuel costs escalate thereafter at three percent, all costs expressed in first-year-of-operation dollars are equivalent to base-year costs. As a result, the first-year COE is also equivalent to the base year COE.

$$\mathrm{COE} = \frac{(\mathrm{CCF})(\mathrm{TOC}) + (\mathrm{OC}_{\mathrm{FIX}}) + (\mathrm{CF})(\mathrm{OC}_{\mathrm{VAR}})}{(\mathrm{CF})(\mathrm{MWh})} \qquad (1)$$

where

COE = revenue received by the generator per net megawatt-hour during the PC plant's first year of operation (but expressed in base-year dollars, \$/MWh or equivalent to mills/kWh)

CCF = capital charge factor, which was 0.124 for high risk IOU, 5-year capital expenditure period

TOC = total overnight cost, expressed in base-year dollars

OC_{FIX} = the sum of all fixed annual operating costs, expressed in base year dollars

OC_{VAR} = the sum of all variable annual operating costs, including fuel at 100% capacity factor, expressed in base year dollars

CF = plant capacity factor, assumed to be constant over the operational period

MWh = annual net megawatt-hours of power generated at 100% capacity factor

3.3 TECHNICAL ASSESSMENT METHODOLOGY AND ASSUMPTIONS

There are several key technical/performance parameters that can also be used to compare the Econamine CO_2 capture process to the sorbent-based CO_2 capture process.

3.3.1 Energy Penalty

For both CO_2 capture processes thermal input (mainly for the regeneration energy) and electricity (for flue gas blowers, CO_2 compression, pumps, etc.) are required for the overall process. For the Econamine process the required electricity and thermal input required was provided in the DOE report.[1] The sorbent-based energy requirements were calculated by the project team based on the conceptual design shown in Figure 2-2. Equation 2 can be used to calculate a non-comprehensive heat duty. Although there are many more terms associated with the actual regeneration energy, only the largest terms are considered in this evaluation. This equation can be used for two objectives: 1) to compare the overall regeneration energy for the two processes and 2) to identify potential energy savings that can be achieved after further research and development of the sorbent-based CO_2 capture process.

$$Q_{regen} = mC_p\Delta T + \Delta H_{rxn} + \Delta H_{vap} \qquad (2)$$

where

Q_{regen} = Regeneration heat duty, Btu/lb CO_2 captured

m = The inverse of the solvent/sorbent delta loading (lb sorbent or solvent per lb CO_2 captured)

C_p = Solution/sorbent specific heat, Btu/lb·°F

ΔT = The temperature difference between adsorption and regeneration

ΔH_{rxn} = The enthalpy of reaction between the amine in solution or on the sorbent, Btu/lb CO_2

ΔH_{vap} = The enthalpy associated with the vaporization of water, Btu/lb CO_2

3.3.2 Environmental Considerations

The objective of the implementation of CO_2 capture is to mitigate the environmental effects related to excessive anthropogenic CO_2 emissions. It is important that the selected CO_2 capture process minimizes all environmental concerns, such as waste and fugitive emissions. There are two potential environmental concerns where the sorbent based CO_2 capture process may offer significant environmental benefits over the Econamine process. One benefit of utilizing a dry sorbent for CO_2 capture is that this material has demonstrated a lower specific heat compared to a solution of MEA in water, which results in a lower amount of sensible heat capacity required to change the temperature of the material. Therefore, it is expected that the water usage for creating the CO_2 capture medium (none for the dry process) and the cooling water requirements will be less. This could be more important at some coal-fired power plants where water resources are increasing in short supply. In addition, the volatile emission of amines has also been considered and is discussed in the Section 4.2.2.

4.0 Results

4.1 ECONOMIC ASSESSMENT AND COMPARISON

In several key areas the engineering firm concluded that whether the basis for the capital cost estimates were provided from either DOE or the engineering firm the sorbent based CO_2 capture resulted in lower cost of electricity increases than the benchmark Econamine based CO_2 capture. The following list was generated to highlight the comparison of the key economic considerations:

- Capital costs, fuel costs and CO_2 TS&M costs of the sorbent-based process were lower than those of the Econamine case, respectively. Specifically, to generate 550 MW_{net} with 90% CO_2 capture, the gross power plant was smaller for when sorbent-based CO_2 capture was used due to the lower thermal input required to operate the CO_2 capture facility.
- The O&M costs of the sorbent-based CO_2 capture were higher than those of the Econamine process, mostly as a result of a high sorbent cost and consumption for an initial fill. With the data that will be collected during 1 MW pilot testing it may be possible to reduce the overall mass of sorbent required in the system and, thus, reduce the initial fill.
- The electric requirements for the compression of CO_2 is significant for both cases. However, because the gross power plant with sorbent-based capture is smaller, less CO_2 must be compressed and the related costs are nominally less. In both CO_2 capture processes blowers are used to overcome the system pressure drop.
- COE/LCOE of Sorbent1 and Sorbent2 cases are lower than those of Econamine1 and Econamine2, respectively, due to a lower capital costs, fuel costs, and CO_2 TS&M costs, offsetting a higher O&M costs.
- Case Sorbent1 resulted in the lowest COE and LCOE, 113.3 and 143.6 mills/kWh, respectively.
- Based on the cost methodology used for calculating cost of electricity, the order of parameters from the largest to the smallest contribution to the COE was:
  For Cases A1 & A2: *Capital* > *Fuel* > *Fixed O&M* > *Variable O&M* > *CO_2 T&SM*
  For Cases B1 & B2: *Capital* > *Fuel* > *Variable O&M* > *Fixed O&M* > *CO_2 T&SM*

In general, a PC plant with the solid sorbent-based CO_2 capture will demonstrate better technical and economic performance than using an aqueous amine capture process as described in the DOE/NETL report. Compared with the PC plant without CO_2 capture published in the DOE Final Report[1], the PC plant with the solid sorbent technology results in a COE increase by 80 percent from 62.9 (escalated to 2011 dollars) to 113.3 mills/kWh. Recall these estimates are based upon a Class IV/V estimate which have an accuracy of +/-50%. Results from pilot testing should be used to accurately size the equipment required for full-scale. The details from the performance and economic comparison are provided in Table 4-1. Although the costs listed in Table 4-1 are greater than the targets identified by the DOE, there are several potential means to significantly lower these costs, which are discussed in Section 5.0.

Table 4-1 Performance and Economic Comparison[*]

Description	Unit	Econamine1 BEC as per DOE	Econamine2 BEC based on experience	Sorbent1 BEC as per DOE	Sorbent2 BEC based on experience
1. Gross Power Output	kW_e	672,700	672,700	656,125	656,125
2. Electrical Output – BP Turbine	kW_e	0	0	47,369	47,369
3. Auxiliary Load Summary	kW_e	122,740	122,740	153,534	153,534
• PC Plant	kW_e	51,550	51,550	50,280	50,280
• SO_2 Polishing Unit	kW_e	0	0	2,230	2,230
• CO_2 Capture Facility	kW_e	22,400	22,400	57,744	57,744
• Compression Unit	kW_e	48,790	48,790	42,858	42,858
4. Net Power Output	kW_e	549,960	549,960	549,960	549,960
5. Thermal Input	kW_{th}	2,102,644	2,102,644	1,865,320	1,865,320
• Gross Output – PC Plant	kW_{th}	1,513,467	1,513,467	1,476,175	1,476,175
• Sorbent Regeneration	kW_{th}	589,177	589,177	389,145	389,145
6. Coal Flow Rate	lb/hr	614,994	614,995	545,581	545,581
7. Coal HHV	Btu/lb	11,666	11,666	11,666	11,666
8. Net Plant HHV Efficiency	%	26.2	26.2	29.5	29.5
9. Net Plant HHV Heat Rate	Btu/kWh	13,046	13,046	11,573	11,573
10. Net kWh @ 85% Capacity Factor	kWh-net	4,095,002,160	4,095,002,160	4,095,002,160	4,095,002,160
11. CO_2 Capture Profile					
• CO_2 Produced	lb/hr	1,459,838	1,459,838	1,295,067	1,295,067
• CO_2 Captured	lb/hr	1,313,960	1,313,960	1,165,561	1,165,561
• CO_2 Emission	lb/hr	145,878	145,878	129,507	129,507
12. Circulating Cooling Water Demand[†]	gpm	540,000	540,000	399,276	399,276
• CW Demand – PC Plant	gpm	200,000	200,000	195,072	195,072
• CW Demand – Capture Facility	gpm	340,000	340,000	204,204	204,204
13. Raw Water Consumption	1000 gal/D	8,081	8,081	6,460	6,460
14. CO_2 Capture Facility					
• Solvent Regeneration Energy	Btu/lb CO_2	1,530	1,530	1,139	1,139
• Total Energy Requirement	MMBtu/hr	2,010	2,010	1,328	1,328
• Steam Extraction Flow	lb/hr	2,219,454	2,219,454	1,429,730	1,429,730
• Amount of Solid Sorbent	lb/hr	N/A	N/A	16,610,220	16,610,220

[†] Assumed maximum circulating CW demands required by the PC plant and the CO_2 capture facility for Cases Econamine1 and Econamine2.

Table 4-1 Performance and Economic Comparison*

Description	Unit	Econamine1 BEC as per DOE		Econamine2 BEC based on experience		Sorbent1 BEC as per DOE		Sorbent2 BEC based on experience	
A. Capital Cost									
1. Bare Erected Cost	$	1,317,434,998		1,797,563,381		1,204,518,214		1,547,066,531	
• PC Plant	$	974,502,282		974,502,282		953,857,228		958,826,649	
• CO_2 Capture Facility	$	302,505,839		782,634,223		215,149,481		552,378,840	
• CO_2 Compression	$	40,426,876		40,426,876		35,511,505		35,861,042	
2. Total Plant Cost	2011$	1,719,001,555		2,431,552,345		1,557,371,701		2,064,504,368	
3. Total Plant Cost	2011$/kW	3,126		4,421		2,832		3,754	
4. Total Overnight Cost	2011$	2,108,905,287		2,981,487,193		1,927,783,145		2,551,290,413	
5. Total Overnight Cost	2011$/kW	3,835		5,421		3,505		4,639	
6. Total As-Spent Cost	2011$	2,404,152,028		3,398,895,400		2,197,672,785		2,908,471,070	
7. Total As-Spent Cost	2011$/kW	4,372		6,180		3,996		5,288	
B. Operating & Maintenance Cost		**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**
1. Total Fixed Operating Costs	$		56,784,795		75,593,565		52,880,649		66,320,549
2. Total Variable Operating Costs	**$**	**2,881,127**	**39,827,356**	**2,881,127**	**47,626,023**	**12,077,712**	**68,138,578**	**12,077,712**	**58,542,353**
• Maintenance Material Cost	$		16,617,203		23,492,742		15,190,044		20,102,994
• Consumables	$	2,881,127	23,210,153	2,881,127	24,133,281	12,077,712	52,948,534	12,077,712	38,439,359
C. Fuel Costs	**$**	**92,862,624**		**92,862,624**		**81,578,386**		**82,381,357**	
D. CO_2 TS&M Costs	$	**25,228,047**		**25,228,047**		**22,160,652**		**22,378,778**	
E. COE contribution (2011$)	**mills/kWh**	**116.3**		**149.2**		**113.3**		**133.3**	
• Capital Costs	mills/kWh	63.9		90.3		58.4		77.3	
• Fixed Operating Costs	mills/kWh	13.9		18.5		12.9		16.2	
• Variable Operating Costs	mills/kWh	9.7		11.6		16.6		14.3	
• Fuel Costs	mills/kWh	22.7		22.7		19.9		20.1	
• CO_2 T&SM Costs	mills/kWh	6.2		6.2		5.4		5.5	
F. LCOE (2011$)	**mills/kWh**	**147.7**		**189.1**		**143.6**		**169.0**	

4.2 TECHNCIAL ASSESSMENT AND COMPARISON

While the overall cost of electricity increase due to CO_2 capture is important, there are other key performance parameters which also must be taken into consideration. Examples of such parameters include energy usage, water requirements, and overall environmental impacts.

4.2.1 Energy Penalty

There are two main sources of energy penalty for the Econamine and sorbent-based CO_2 capture processes, electricity (i.e., parasitic power consumption) and thermal (steam). In both cases electricity is used for flue gas blowers (to overcome the absorber/adsorber pressure drop), CO_2 compression, and cooling water pumps. When the engineering firm compared the overall energy penalty for the two processes, they concluded that the sorbent-based process B resulted in a lower energy penalty. Specifically, they found that the net plant efficiency when solid sorbent CO_2 capture is employed is 29.5 percent, which is a 12.6 percent improvement compared to the benchmark Econamine technology.

To understand why the thermal input for the sorbent-based CO_2 capture process is lower than that of the Econamine process, it is helpful to consider the regeneration heat duty (Q_{regen}) defined in equation 2. Based on the DOE report the total Econamine Q_{regen} is 1530 Btu/lb CO_2 captured. On the right hand side of equation 2 are the key terms contributing to the Q_{regen}. The first term is related to the heat input required to increase the sorbent or solvent from the adsorption temperature to the regeneration temperature. The term m is the inverse of the delta CO_2 loading. Based on mass balances provided in previous DOE reports reported a delta CO_2 loading for aqueous MEA system to be approximately 4.5 lb CO_2/100 lb fresh solvent.[1] It is assumed that the specific heat of for aqueous MEA is approximately the same as that of pure water, which is 1.0 Btu/lb·°F. The absorption and regeneration temperatures of the Econamine process are approximately 122°F and 243°F, respectively.[4] Therefore, the difference between the absorption temperature and the regeneration temperature (ΔT) is 120°F. The sensible heat input for the aqueous MEA process is calculated, it is (100 lb fresh solvent/4.5 lb CO_2)*(1.0 Btu/lb·°F)*(120°F) = 2670 Btu/lb CO_2 captured. However, ΔT in equation 2 is only the temperature where active heat input is required to cause a temperature change. The Econamine process employs a lean/rich heat exchanger to reduce the energy requirement due to sensible heat. Most of the sensible heat from the lean solvent leaving the regenerator is transferred to the rich sorbent before it enters the regenerator. Based on public literature, it is accepted that this lean/rich heat exchanger can be operated at an approach temperature (the temperature difference between the lean solvent exiting the heat exchanger and the rich solvent entering it) of approximately 18°F.[3] The sensible heat requirement for the Econamine process is then calculated to be (100 lb fresh solvent/4.5 lb CO_2)*(1.0 Btu/lb·°F)*(18°F) = 400 Btu/lb CO_2 captured. Insufficient information was provided in the DOE report to calculate the exact values of the second and third terms in equation 2. However, using estimates from public literature a general estimate of these terms was developed. The second term on the right hand side of equation 2, ΔH_{rxn}, is the enthalpy generated due to the exothermic reaction with CO_2. For the Econamine process it is assumed to be 700 Btu/lb CO_2 captured.[5] The last term on the right hand side of equation 2 is related to the vaporization of water. From literature, this value for aqueous MEA is approximately 290 Btu/lb CO2 captured.[5] Therefore, the total Q_{regen} ~ 1400 Btu/lb CO_2, which is less than the reported 1530 Btu/lb CO_2 (meaning that additional heat duty requirements should be taken into account or our calculations were not conservative enough).[2] If Q_{regen} is calculated without the lean/rich heat exchanger the total energy penalty is 2670 + 700 + 290 = 3660 Btu/lb CO_2. Therefore, the presence of the lean/rich heat exchanger reduced the overall regeneration energy duty by approximately 60%.

For solid sorbents with no heat recovery and no lean/rich heat exchanger, the sensible heat can be calculated similarly, (100 lb fresh sorbent/7.0 lb CO_2)*(0.25 Btu/lb·°F)*(144°F) = 514 Btu/lb CO_2. The sensible heat duty for the sorbent based process is 80% lower than that of the aqueous amine process without heat recovery. The enthalpy of reaction (ΔH_{rxn}), which will be confirmed through 1 MW pilot testing, is projected to be approximately 587 Btu/lb CO_2.[4] In the past the enthalpy of vaporization was expected to be negligible for dry sorbents. However, different sorbents will exhibit varied behavior regarding moisture uptake. While some sorbents may be hydrophobic, the sorbent used as the basis to complete the cost and technical assessment will remove moisture from the flue gas and release it in the regenerator. On a mass basis the moisture removed from the flue gas is signficantly less than that of the CO_2, but the enthalpy of vaporization term (ΔH_{vap}) can not be neglected in the energy balance. Using laboratory data generated to date, the expected delta loading for moisture on the sorbent will be approximately 0.9 lb H_2O/lb fresh sorbent. Assuming that the enthalpy for the release of moisture from the sorbent is the same as that the vaporization of water, ~970 Btu/lb H_2O,[6] this can be converted to 235 Btu/lb CO_2 . Therefore, the regeneration heat duty for the sorbent-based CO_2 capture process is 514 + 587 + 235 = 1340 Btu/lb CO_2. Even without the presence of heat recovery the CO_2 capture process offers a lower regeneration heat duty. If it is deemed to be economically feasible to implement heat recovery for the sorbent-based process then the regeneration heat duty can be reduced. It is important to note that the heat duty calculated for the sorbent-based CO_2 capture process is based entirely on laboratory experiments and public literature. Operation of the 1 MW pilot is extremely important to determining the actual values of the heat duty and other key operating paramters. A series of tests will be performed during pilot testing to quantify the latent heat of reaction for both CO_2 and H_2O adsorption. Clearly, this information is critical to accurately calculate the energy requirements for solid sorbent based CO_2 capture.

The thermal input requirements for the Econamine and the sorbent-based processes are provided in Figure 3-1. As expected, due to the lower regeneration heat duty the overall thermal input is lower for the sorbent based process.

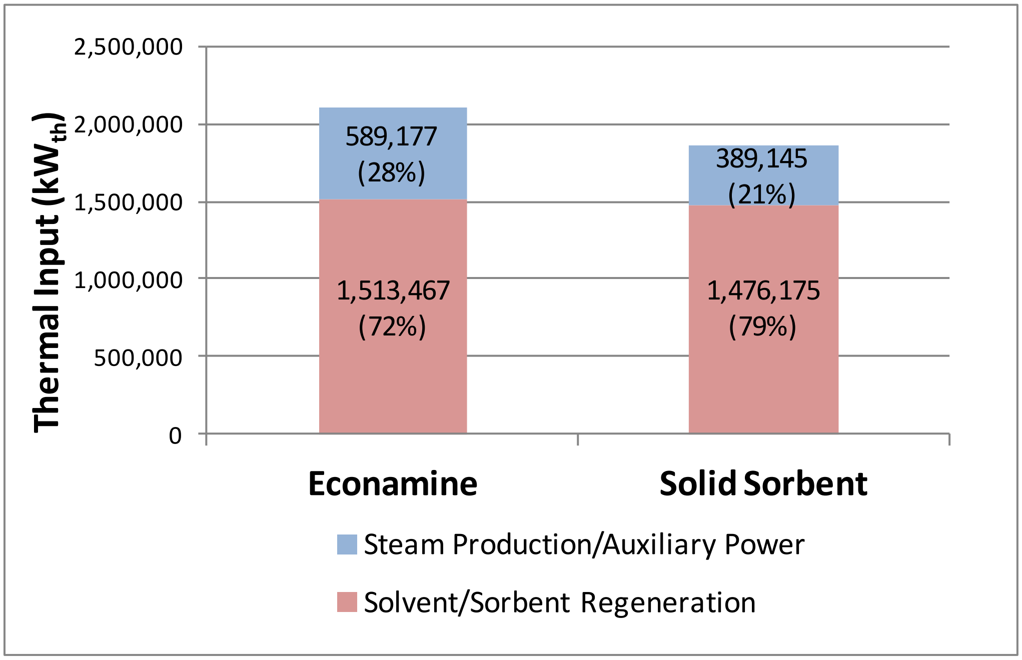

Figure 3-1 Thermal Input

4.2.2 Environmental Considerations

The coal supply to the PC plant is linearly related to the power plant gross thermal output. Because the gross thermal output for the power plant with sorbent-based CO_2 capture is tr percent less than that of the Econamine process less coal is burned and thus, less mercury (Hg), SO_2, SO_3 (sulfur trioxide), NO_x, etc. will be created. This beneficial environmental impact must also be taken into consideration when comparing the Econamine and the sorbent based processes.

Similarly, the raw water consumption of the solid based process (~6.5M gallon/D) is 20 percent lower than that of the Econamine process (~8.1M gallon/D). This can be attributed to the fact that the gross power plant is smaller when the sorbent-based CO_2 capture process is implemented. Because water is likely to become an increasingly scarce resource in the future, less water consumption may be an important property for the sorbent based process. The calculation methodology from the DOE report[1] is used to produce water use and costs. Results on the raw water consumption are shown in Figure 4-1. As per the DOE report[1], it was assumed that 50 percent of the raw water is sourced from groundwater at no cost.

Figure 4-1 Raw Water Utilization

Another significant environmental concern is related to the volatile emissions of amines. At the expected regeneration temperatures the vapor pressure of 3.5 M aqueous MEA is approximately 0.0725 psi. Although the sorbent has been degraded in the laboratory using excessively high temperatures, no measureable volatile emissions were recorded using a mass spectrometer. It is possible that the amine on the selected sorbent remains covalently attached to the surface of the substrate even when being converted to urea, heat stable salts, or being otherwise degraded.

Another environmental consideration is related to potential spills. If the proposed sorbent were to spill either in transport or at the power plan, unlike an aqueous solution the sorbent will not seep into the soil. The sorbent could more easily be cleaned up without serious environmental consequences. The sorbent is not considered hazardous by the OSHA Hazard Communication Standard, thus waste disposal should be in accordance with the plants locale environmental controls.

5.0 Discussion and Conclusions

Two potential post-combustion CO_2 capture processes have been compared; the Econamine process employs aqueous MEA and the other process is a process based on an aminated sorbent under development. The two options were compared with respect to cost, energy penalty, and environmental performance.

The sorbent-based CO_2 capture process resulted in a lower energy penalty and lower overall increase in the cost of electricity. However, the improvements versus aqueous MEA are not enough to meet the DOE's COE goals. With further improvements it is possible that the sorbent-based CO_2 capture process may be able to make significant progress towards the DOE goals. Potential areas of improvement include:

- Heat integration
    - Heat recovery between the rich and lean sorbent – this must be accomplished in such a way that the energy recovery is significant enough to offset the added capital costs for the heat exchangers
    - Integration within the power plant – using heat from the power plant or the CO_2 compression system to reduce the overall energy penalty
- Reduce capital costs – for the preliminary conceptual design discussed in this report several key assumptions were made. It is possible with the data from the 1 MW pilot that the overall process size and complexity could be reduced, which would importantly result in a decrease in the overall capital costs. A reduction in the number of necessary trains would result in a significant cost reduction.
- Improve sorbent performance and characteristics
    - Working CO_2 capacity
    - Attrition resistance
    - Cost
    - Thermal conductivity
    - Moisture loading
    - Etc.
- Optimize steam condition at IP/LP crossover to eliminate a need of the BP turbine and to integrate into the steam cycle in a more efficient manner.

A unique process for CO_2 capture using solid sorbents has been developed. In the absence of large scale (i.e. pilot-scale) testing to date, it was necessary to make many different assumptions regarding the equipment and sorbent performance under the expected operating conditions. To fully capture the true range of operation and fully assess synergistic effects of the entire system operation it is important to continue this work to the 1 MW pilot scale to validate the following key performance parameters:

- CO_2 delta loading
- H_2O delta loading

- Enthalpy of reaction/adsorption of CO_2
- Enthalpy of reaction/adsorption of H_2O
- Reaction kinetics
    - CO_2 uptake limited by kinetics in the adsorber
    - CO_2 release limited by mass diffusion in the regenerator
- Effective heat transfer coefficients
    - Cooling water requirements
    - Steam requirements
- CO_2 partial pressure in the regenerator
- Reactor pressure drop
- Physical attrition rate
- Successful isothermal operation
- Optimize process variables:
    - Temperature
    - Sorbent Circulation Rate
- Identify sorbent bed height required to maintain 90% CO_2 capture

Using results from the 1 MW pilot test a more in depth techno-economic assessment can be completed. If the process will not meet the DOE's goals related to the cost of electricity increase as described in this report, sorbent properties and process conditions that would result in an acceptable cost increase will be identified.

6.0 List of Abbreviations

AACE	Association for the Advancement of Cost Engineering
BACT	best available control technology
BEC	bare erected cost
Btu/lb	British thermal unit per pound
BP	back pressure
°C	degree Celsius
CCF	capital charge factor
CF	capacity factor
CO_2	carbon dioxide
COE	cost of electricity
C_p	specific heat
DOE	Department of Energy
Econamine	Fluor Econamine FG Plus^{SM}
EPC	engineering, procurement, and construction
EPCM	engineering, procurement, construction, and management
°F	degree Fahrenheit
FGD	flue gas desulfurization
ft/s	foot per second
gpm	gallons per minute
KO	knockout
kW_e	kilowatt of electrical energy
kW_{th}	kilowatt of thermal energy
ΔH_{rxn}	enthalpy of reaction
ΔH_{vap}	enthalpy associated with the vaporization of water
H_2SO_4	sulfuric acid
H_2O	water
Hg	mercury
HSS	heat stable salts
IOU	investor-owned utility
IRROE	internal rate of return on equity
IP	intermediate pressure
lb/ft^3	pound per foot cubed
LCOE	Levelized cost of electricity
LNBs	low NO_x burners
LP	low pressure
M	million
m	inverse of the solvent/sorbent delta loading
MEA	monoethanolamine
mills/kWh	milliseconds per kilowatt hour
MW	megawatt
MWh	megawatt-hour
MU	makeup
NaOH	sodium hydroxide
NETL	National Energy Technology Laboratory

NO_x	nitrogen oxides
OC	operating costs
O&M	operations and maintenance
OFA	over-fire air
OSHA	Occupational Safety and Health Administration
PC	pulverized coal
PM	particulate matter
psi	pounds per square inch
psia	pounds per square inch absolute
Q_{regen}	regeneration heat duty
ΔT	temperature difference between adsorption and regeneration
TASC	total as-spent cost
TOC	total overnight cost
TPC	total plant cost
TS&M	transport, storage & monitoring
SCR	selective catalytic reduction
SO_2	sulfur dioxide
SO_3	sulfur trioxide
U.S.	United States
WT	water treatment

7.0 References

1. Final report – Carbon Dioxide Capture from Existing Coal-Fired Power Plants, DOE/NETL- 401/110907, November 2007.

2. Final report – Cost and Performance Baseline for Fossil Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity, DOE/NETL-2007/1281, August 2007.

3. Fisher, Kevin S., Carrie Beitler, Curtis Rueter, Katherine Searcy, Dr. Gary Rochelle, Majeed Jassim, and Jose Figueroa. “Integrating MEA Regeneration with CO_2 Compression to Reduce CO_2 Capture Costs”, Paper Presented at the Fourth Annual Conference on Carbon Capture & Sequestration, Alexandria, VA, May, 2005.

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Techno-Economic Report Commercial – Scale Solid Sorbent CO2 Capture Process: 550 MW Pulverized Coal-Fired Power Plant

Updated Report Based on Solid Sorbent Pilot Data

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Sign-off Sheet

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A	Dec 4, 2015	Draft	BT/TS		MR	
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Executive Summary

The following report provides a Techno-Economic Assessment (TEA) of a generic 550 MW (net) subcritical Pulverized Coal-fired (PC) power station with integration of a carbon capture facility based on a solid sorbent technology. The objective of this report is to determine the economic viability of the emerging solid sorbent technology by comparing it to an established absorption technology at full-scale commercialization. In this study, the benchmark technology selected for comparison is the Fluor Econamine FG Plus^{SM} (Econamine) process.

Previously, a TEA based on preliminary laboratory performance was produced by Stantec[2]. Since then, the technology has been piloted at a 1 MW equivalent scale. This updated TEA includes performance factors obtained from this pilot.

Table E-1 compares both cases studied. As can be seen, solid sorbents are not competitive with the amine based economic case set forth by the DOE report entitled "Cost and Performance Baseline for Fossil Energy Plants – Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 3"[1]. The Levelized Cost of Energy (LCOE) for solid sorbents is approximately 2.2 times higher than the amine case from the report.

The major causes for poor results are due to low working capacity of the sorbent discovered during piloting of the. A low working capacity means that the difference in loading between rich and lean is reduced, thus the amount of sorbent required for 90% capture of emitted CO_2 grows. This leads to multiple equipment trains for absorption and regeneration, resulting in very high capital costs.

Further development of the sorbent, perhaps reformulating to increase working capacity, could potentially restore competitiveness of this technology.

Table E-1 Performance and Economic Comparison in 2015 Dollars

Description	Unit	Amine Technology	Solid Sorbent Technology
Performance			
1. Gross Power Output	kW_e	672,700	718,738
2. Electrical Output – BP Turbine	kW_e	0	32,525
3. Auxiliary Load	kW_e	122,740	201,293
4. Net Power Output	kW_e	549,960	549,970
5. Thermal Input	kW_{th}	2,102,644	2,383,129
6. Coal Flow Rate	lb/hr	614,994	697,032
7. Net Plant HHV Efficiency	%	26.2	23.1
8. Net Plant HHV Heat Rate	Btu/kWh	13,046	14,785
9. CO_2 Capture Profile			
• CO_2 Produced	lb/hr	1,459,838	1,651,831
• CO_2 Captured	lb/hr	1,313,960	1,486,648
• CO_2 Emission	lb/hr	145,878	165,183
10. Circulating CW Demand	gpm	540,000	487,120
11. Raw Water Consumption	1000 gal/D	8,081	7,802
12. Carbon Capture Facility			
• Regeneration Energy	Btu/lb CO_2	1,530	2,042
• Total Energy Requirement	MMBtu/hr	2,010	3,036
• Steam Extraction Flow	lb/hr	1,931,497	2,709,721
			Costs

A. Capital Costs			
1. Bare Erected Costs	$	1,415,831,000	3,377,994,721
2. Total Plant Costs	$	1,847,390,000	4,724,953,270
3. Total Overnight Costs	$	2,266,414,000	5,991,041,269
4. Total As-Spent Costs	$	2,583,712,000	6,355,138,125
B. O&M Costs			
• Total Fixed Initial Costs	$	N/A	N/A
• Total Fixed Annual Costs	$	61,025,905	141,311,446
• Total Variable Operating Initial Costs	$	3,096,311	180,802,311
• Total Variable Operating Annual Costs	$	42,801,959	132,049,027
C. Fuel Costs	$	99,798,295	113,110,975
D. CO_2 TS&M Costs	$	27,112,265	30,675,509
E. COE contribution (2015$)	mills/kWh	125.0	283.9
• Capital Costs	mills/kWh	68.6	181.4
• Fixed Operating Costs	mills/kWh	14.9	34.5
• Variable Operating Costs	mills/kWh	10.5	32.9
• Fuel Costs	mills/kWh	24.4	27.6
• CO_2 T&SM Costs	mills/kWh	6.6	7.5
F. LCOE (2015$)	mills/kWh	158.7	359.9

The cost data is repeated in February 2011 dollars, below:

Table E-1a Economic Comparison in 2011 Dollars

A. Capital Costs			
1. Bare Erected Costs	$	1,317,435,000	3,143,235,000
2. Total Plant Costs	$	1,719,002,000	4,396,584,000
3. Total Overnight Costs	$	2,108,905,000	5,574,683,000
4. Total As-Spent Costs	$	2,404,152,000	6,355,138,000
B. O&M Costs			
• Total Fixed Initial Costs	$	N/A	N/A
• Total Fixed Annual Costs	$	56,785,000	131,491,000
• Total Variable Operating Initial Costs	$	2,881,000	168,237,000
• Total Variable Operating Annual Costs	$	39,827,000	125,219,000
C. Fuel Costs	$	92,863,000	105,250,000
D. CO_2 TS&M Costs	$	25,228,000	28,544,000
E. COE contribution (2011$)	mills/kWh	116.3	264.2
• Capital Costs	mills/kWh	63.9	168.8
• Fixed Operating Costs	mills/kWh	13.9	32.1
• Variable Operating Costs	mills/kWh	9.7	30.6
• Fuel Costs	mills/kWh	22.7	25.7
• CO_2 T&SM Costs	mills/kWh	6.2	7.0
F. LCOE (2011$)	mills/kWh	147.7	334.9

The cost per ton of carbon dioxide captured is presented below in Table E-2. It is calculated using the LCOE established previously, compared against the amount of CO_2 captured each year.

Table E-2: $/ton CO_2 Captured Results

	Case A	Case B
2015 Dollars	71 $/ton	272 $/ton
2011 Dollars	66 $/ton	253 $/ton

1.0 INTRODUCTION

The following report provides details on the preliminary technical and economic performance of a generic 550 MW (net) subcritical Pulverized Coal-fired (PC) power station with integration of a carbon capture facility using a conventional amine-based absorption process (Case A), and a novel solid sorbent-based process (Case B), to remove carbon dioxide (CO_2) from the flue gas stream. The objective of this report is to determine the economic viability of the emerging solid sorbent technology by comparing it to an established absorption technology at full-scale commercialization. In this study, the benchmark technology selected for comparison is the Fluor Econamine FG Plus^{SM} (Econamine) process.

Case studies on a 550 MW (net) PC plant and its integration with the Econamine process (Cases 9 and 10) previously conducted by DOE/NETL, and published in the Final Report entitled "Cost and Performance Baseline for Fossil Energy Plants – Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 3," provide the starting point for the cases in this study. Full descriptions of the generic PC plant and the Econamine process can be found in the DOE/NETL final report[1].

Case 10 from the DOE/NETL final report[1] is used to represent Case A and form the basis for Case B, while Case 9 is used for comparison purposes. To make a fair comparison, process configuration/conditions (except flow rate) of peripheral units outside the carbon capture battery limits for Case B are assumed to remain identical to those of Case A, unless some modifications are required to meet the requirement of the solid sorbent-based process.

Prior to the carbon capture facility, in general, the PC plant is equipped with the Best Available Control Technology (BACT) to comply with environmental targets. The Flue Gas Desulfurization (FGD) system used for sulfur dioxide (SO_2) removal is a wet limestone forced oxidation absorber, which will produce gypsum as a byproduct. A fabric filter is used to control Particulate Matter (PM). The nitrogen oxides (NO_x) concentration is reduced by using a combination of Low NO_x Burners (LNBs) and Over-Fire Air (OFA) systems in the boiler and installing Selective Catalytic Reduction (SCR) prior to the air heater.

Since both cases are evaluated on a common 550 net output basis, the gross steam turbine output must vary to account for different auxiliary and heat power demand for the two cases. As a result, the PC plant size and the amount of coal required will vary accordingly to ensure a constant net output of 550 MW. Thermal input generated by burning coal must be sufficient to supply an electricity demand for both PC plant and carbon capture facility as well as energy requirement for solvent / sorbent regeneration.

1.1 SUBCRITICAL PULVERIZED COAL-FIRED (PC) POWER PLANT

This report focuses on 550 MW net output steam-power cycle operating under subcritical steam conditions uses a single reheat with 2,415 psia and 1,050°F main steam, and 565.5 psia and

1,050°F reheat steam. The steam turbines are composed of a high pressure (HP) turbine, an intermediate pressure (IP) turbine and a low pressure (LP) turbine. Steam leaving the HP turbine returns to the reheater in the boiler where it is reheated and expanded to IP and LP turbines and eventually a condenser. The condenser condenses the steam and rejects heat to the cooling tower. Condensate accumulates in the hotwell and is pumped by the condensate extraction pump through a gland seal condenser, four low pressure (LP) feedwater heaters, and the deaerator. The boiler feed pump underneath of the deaerator pressures the condensate through two HP feedwater heaters.

Bled steam at the last stage of the IP turbine is extracted to mechanically drive the boiler feed turbine which is connected to the boiler feed pump. The condensate then enters the economizer and boiler drum. The drum separates vapor from liquid. Vapor is superheated by the superheater and is sent to the HP turbine. Meanwhile, liquid from the boiler drum enters the water tubes, attached to boiler inside wall. Heat from the boiler is transferred through these tubes to boil water to steam.

The performance of the unit is based on Illinois no. 6 bituminous coal. Flue gas from coal combustion transfers energy to a number of tubes inside the boiler. The arrangement of the boiler tubes is based on conventional boiler configuration composed of a superheater, reheater, boiler drum, water wall tubes and economizer. A tri-sectional air preheater is the last equipment to pick up heat from the flue gas to the inlet air of primary and forced draft air.

1.2 CARBON CAPTURE FACILITY

As mentioned previously, two cases are investigated: a conventional absorption process and the solid sorbent-based process. Both are designed to capture 90 percent of the CO_2 from the whole flue gas stream generated by the 550 MW (net) PC plant.

Case A: Conventional Absorption Process

This process is based on the Fluor Economine FG Plus^{SM} technology using formulated aqueous monoethanolamine (MEA) solution with a proprietary corrosion inhibitor as a solvent to absorb CO_2 from the flue gas stream. To minimize the formation of Heat Stable Salts (HSS), an SO_2 polishing unit using 20wt percent sodium hydroxide (NaOH) must reduce the SO_2 concentration in the flue gas stream exiting the FGD down to 10 ppmv or less. The flue gas exiting the SO_2 polishing is pressurized by a fan to overcome a 2-psi pressure drop across the absorber.

The flue gas, with a temperature of approximately 135°F, then enters at the CO_2 capture equipment and flows counter-currently against the lean MEA-based solution introduced at the top of the absorber. After removing CO_2 the rich MEA-based solution is collected at the absorber bottom, whereas the CO_2-free flue gas is released at the absorber top. The CO_2 rich solution is preheated by the CO_2 lean solution from the solvent stripper in the lean/rich cross exchanger. The cooled CO_2 lean solution is directed to the absorber for a complete circulation, whereas the heated rich solution is directed to the solvent stripper for CO_2 stripping and solvent regeneration using steam from the IP/LP crossover pipe of steam turbine. For sequestration purpose, the CO_2 product gas is compressed to 2,215 psia by a six-stage compressor and dehydrated to 40°F with triethylene glycol.

Case B: Solid Sorbent-Based Process

The solid sorbent-based process employs an amine-impregnated solid sorbent technology developed by ADA-ES. Previously, a TEA based on preliminary laboratory performance was produced by Stantec[2]. Since then, the technology has been piloted at a 1 MW equivalent scale. This updated TEA includes performance factors obtained from the pilot, which is presented below:

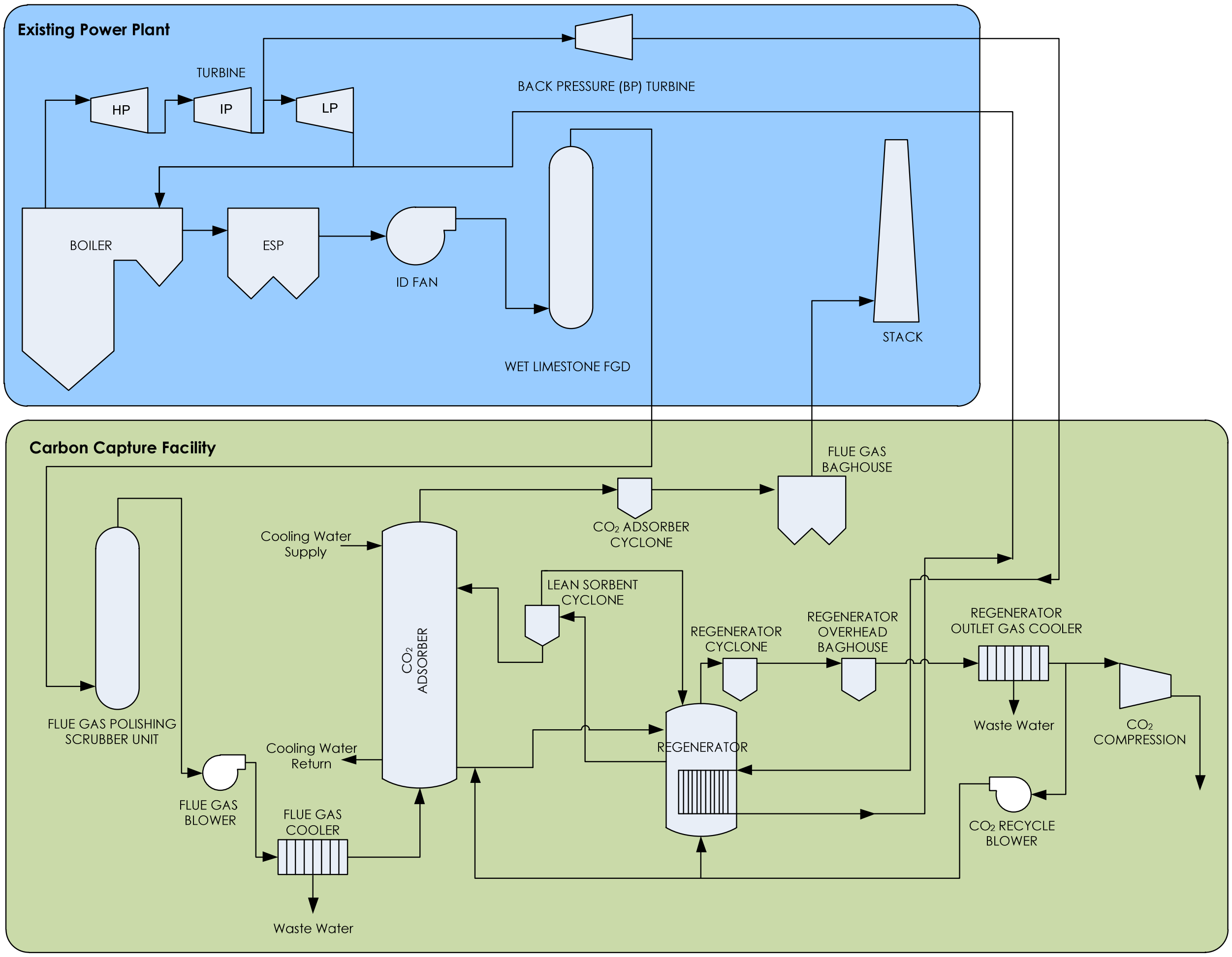
- Particles physically resemble polystyrene beads with respect to size. Particles were milled to a mean particle diameter of about 165 – 180 µm.
- Particle density is 36.6 lb/ft^3.
- Sorbent working capacity is 5.4 %wt (lb CO_2/100 lb sorbent). This corresponds to 65 minutes residence time in the adsorber.
- Average bed density in the adsorber is 23.15 lb/ft^3 for the bottom bed, 24.32 lb/ft^3 for the middle bed and 23.34 lb/ft^3 for the top bed. High bed density results from reduced gas velocity through the adsorber to approach 90% CO_2 removal efficiency.
- Average bed height in the adsorber is 7.99 ft for the bottom bed, 6.00 ft for the middle bed and 17.84 ft for the top bed.
- Superficial gas velocity of the flue gas entering the adsorber is 0.7 ft/s.
- Thermal energy for regeneration is 2042 Btu/lb CO_2 adsorbed.
- Adsorption temperature is 104°F.
- Regeneration temperature is 248°F.
- Flue gas temperature after the SO_2 polishing unit is 135°F.

Section 2.0 provides more information on the integration of this process into the PC plant along with a detailed process description.

2.0 SOLID SORBENT-BASED PROCESS

The concept of a solid-based CO_2 capture process integrated into the PC plant is demonstrated in Figure 2-1. Figure 2-2 shows a conceptual representation of adsorption and desorption equipment. Full Process Flow Diagram for the 1 MW solid sorbent CO_2 capture pilot plant can be found in Appendix A.

Figure 2-1 Process Flow Diagram: Solid Sorbent Technology



Flue gas is routed from the existing plant through ductwork to the new carbon capture facility. The SO_2 concentration in the flue gas before entering the carbon capture facility must be reduced to low levels by a similar SO_2 polishing unit used in the Econamine process. A flue gas blower is installed to provide sufficient pressure to pass the flue gas through the carbon capture equipment, and return it via ductwork back to the original stack. As a result, the pressure of the flue gas is boosted by approximately 7 psi, causing the flue gas temperature to rise from 135°F to

219°F. The hot wet flue gas is then directed to the flue gas cooler, where the flue gas is cooled down to 104°F by exchanging heat with cooling water and the water Knockout (KO) drum where water vapor is separated from the flue gas.

The cool flue gas then enters an isothermal three-stage fluidized bed adsorber. The stages are separated by horizontal perforated trays, which allow the flue gas entering at the bottom to flow through the adsorber and fluidize the sorbent bed in a churn flow regime. The treated flue gas is released at the adsorber top and routed to a baghouse, which removes any entrained sorbent ensuring none is lost or emitted as a pollutant when the flue gas is returned to the stack. The baghouse will remove very fine sorbent that may have been damaged through attrition, thus collected solids could be removed for disposal if required. In addition, the adsorber is equipped with a bundle of serpentine tubes in each stage to maintain isothermal operation by exchanging exothermic heat generated from chemical reaction between CO_2 and amine sorbent with cooling water.

To maximize the driving force for mass transfer, this adsorber simulates continuous counter-current operation by installing a standpipe between each stage. The standpipe functions as a passage for the sorbent to cascade down from the top to the bottom of the adsorber, while adsorbing CO_2 from the upward flowing flue gas. As the sorbent rich in CO_2 is withdrawn at the bottom, lean sorbent is simultaneously added at the top of the adsorber to replenish the bed. The amount of the rich sorbent discharged is controlled by a sorbent slide valve and pneumatically conveyed with CO_2 to the regenerator via a rich sorbent riser.

In an isothermal single-stage fluidized bed regenerator, the rich sorbent releases CO_2 by means of heat from steam circulating in an internal heat exchanger. Since the temperature of steam (306°F) extracted from IP/LP crossover section of the PC plant steam turbine is hotter than the regeneration temperature of the solid sorbent (248°F), the expansion of the steam through a Back Pressure (BP) turbine before entering the exchanger is required to prevent the thermal degradation of the amines impregnated in the solid sorbent and also to generate additional electricity. Saturated steam at 45 psia is obtained at the BP turbine exit.

After being directed through a baghouse downstream of the regenerator, a portion of the CO_2 is routed to the CO_2 recycle compressor, while the remaining CO_2 is sent to the compression plant similar to the Econamine process. The recycle compressor provides enough pressure for the CO_2 to promote adequate fluidization in the regenerator and to convey sorbent from the adsorber to the regenerator in the rich sorbent riser. The compression plant features wet stages, followed by dehydration, then dry stages to compress the CO_2 to supercritical conditions.

To complete the cycle, the hot lean sorbent from the regenerator is transferred via a lean sorbent riser, using the conveying media of CO_2 lean flue gas. The top bed of the adsorber is deeper to accommodate more heat exchange area required to cool the sorbent down to the adsorption temperature. Originally, 4 absorption / regeneration trains were sufficient to treat the incoming flue gas (based on laboratory results for sorbent performance). The pilot plant results have indicated that in order to provide sufficient residence time for 90% capture, 44 parallel trains are now required.

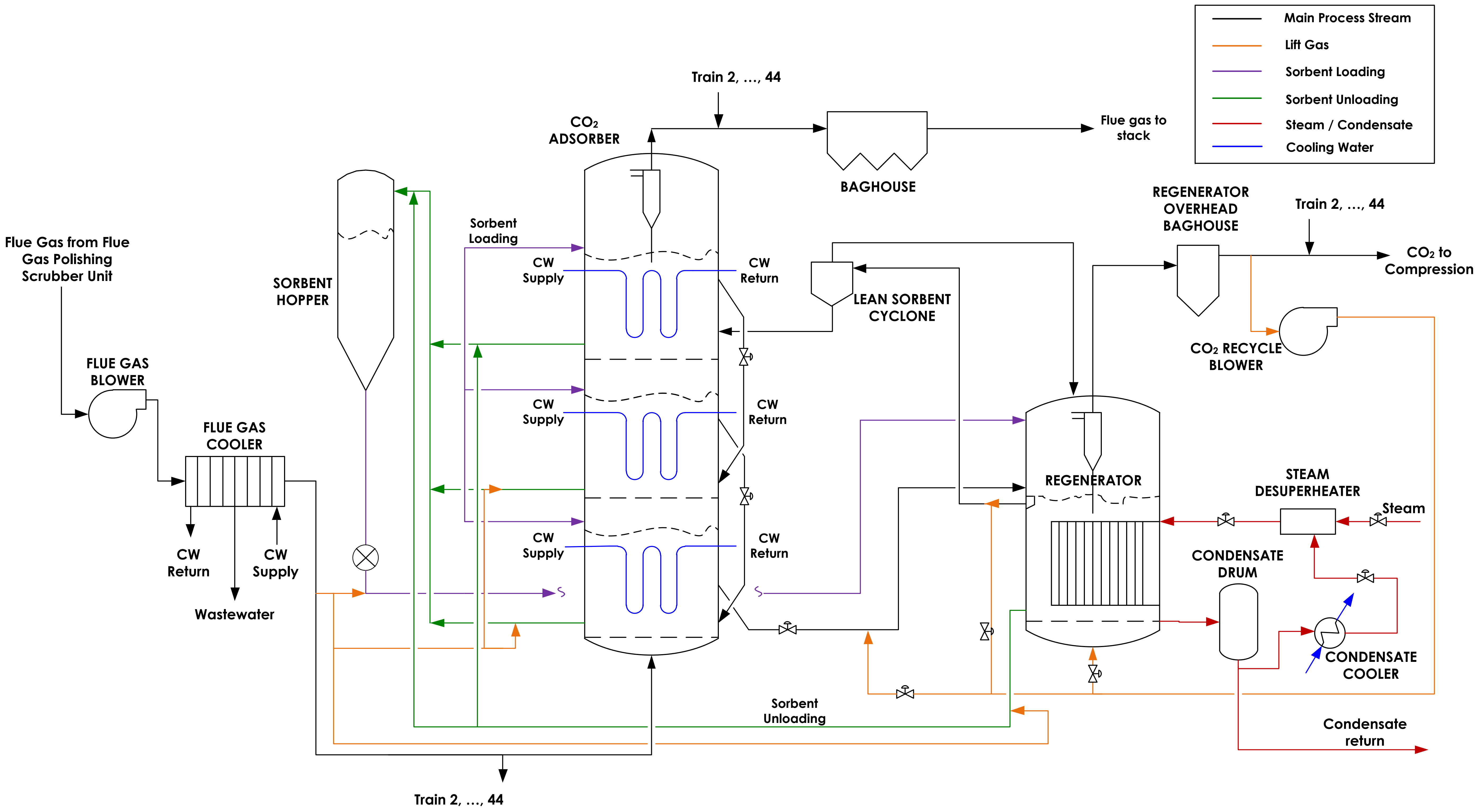


Figure 2-2 Process Flow Diagram of the Solid–Based CO_2 Capture Process

3.0 PROCESS SIMULATION

3.1 PROCESS VALIDATION

This study employed GateCycle, a thermal system designing software, to model a pulverized boiler / furnace, steam cycle and heat rejection system. A process validation with the DOE's Cases 9 and Case 10 was conducted in order to gain necessary confidence to proceed on the process integration with the solid sorbent-based process. The results showed a good agreement between the DOE's cases and the GateCycle process simulation models. Full details can be found in Appendix B.

3.2 ADA-ES CO_2 ADSORBENT PROCESS INTEGRATION

Thermal input supplied by coal combustion is utilized in two different ways: providing both the gross output required to operate the 550 MW (net) PC plant, and the thermal regeneration energy requirement for the sorbent system.

3.2.1 Gross Power Output

The gross power output varies with auxiliary load of both the PC plant and the solid sorbent-based carbon capture facility and electrical output produced from the BP turbine. The auxiliary load of the PC plant and the BP turbine output are obtained by GateCycle simulation whereas that of the capture facility is obtained by linearly scaling the estimated auxiliary load with CO_2 removal rate.

For CO_2 compression, ProMax chemical process simulation is used to estimate the auxiliary loads required to compress the captured CO_2 at a given set of carbon capture process conditions. A relationship between the CO_2 product flow rate and CO_2 compression auxiliary load obtained from the ProMax simulation was used as an input to GateCycle for total gross power output calculation.

A series of iterations in GateCycle is programmed to vary input parameters to meet a net power output of 550 MW. All process parameters (e.g., coal consumption, flue gas flow, BP output and steam extraction flow rate) and auxiliary loads for the PC plant and its integration with the carbon capture facility are constantly changing as the GateCycle progresses to the 550 MW net output target. The final outputs are obtained when the model converges. This model output can also be found in Appendix B.

3.2.2 Steam Consumption

Steam is extracted at the last stage of the IP turbine to the BP turbine. The BP turbine controls the steam pressure to 45 psia before entering the reboilers in the solid sorbent-based carbon capture facility. A rate of extracted steam to the reboiler depends on the CO_2 removal flow rate at a given thermal regeneration energy of 2,042 Btu/lb CO_2 obtained from the actual pilot plant data.

3.2.3 Performance

Performance of the carbon capture facility using the amine technology and the solid sorbent-based technology is summarized in Table 3-1.

Table 3-1 Performance Comparison

Parameter	Unit	Amine Technology	Solid Sorbent Technology
Gross Power Output	**kW$_e$**	**672,700**	**718,738**
Back Pressure Turbine Power Output	**kW$_e$**	**-**	**32,525**
Total Gross Power Output	**kW$_e$**	**672,700**	**751,263**
Auxiliary Load Summary	**kW$_e$**	**122,740**	**201,293**
• PC Plant			
Coal Handling and Conveying	kW$_e$	540	612
Pulverizers	kW$_e$	4,180	4,738
Sorbent Handling & Reagent Preparation	kW$_e$	1,370	1,553
Ash Handling	kW$_e$	800	907
Primary Air Fans	kW$_e$	1,960	2,222
Forced Draft Fans	kW$_e$	2,500	2,833
Induced Draft Fans	kW$_e$	12,080	13,671
SCR	kW$_e$	70	79
Baghouse	kW$_e$	100	113
Wet FGD	kW$_e$	4,470	5,066
Miscellaneous Balance of Plant[1,2]	kW$_e$	2,000	2,000
Steam Turbine Auxiliaries	kW$_e$	400	400
Condensate Pumps	kW$_e$	700	621
Circulating Water Pumps	kW$_e$	11,190	3,703
Ground Water Pumps	kW$_e$	1,020	775
Cooling Tower Fans	kW$_e$	5,820	1,918
Transformer Losses	kW$_e$	2,350	2,511
Sub Total	*kW$_e$*	*51,550*	*43,722*

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[1] Turbine driven boiler feed pumps are used.

[2] Plant control systems, lighting, HVAC and miscellaneous low voltage loads are included.

Parameter	Unit	Amine Technology	Solid Sorbent Technology
• Flue Gas Polishing Scrubber Unit			
Recirculating Pumps	kW_e	0	2,276
Bleed Pumps	kW_e	0	569
Sub Total	*kW_e*	*0*	*2,845*
• Carbon Capture Facility[3]			
Circulating Water Pump	kW_e	N/A	4,802
Cooling Tower Fan	kW_e	N/A	2,503
Condensate Pump	kW_e	N/A	325
Flue Gas Compressor	kW_e	N/A	53,617
CO_2 Recycle Compressor	kW_e	N/A	14,787
Lean Sorbent Lift Compressor	kW_e	N/A	16,527
Sub Total	*kW_e*	*22,400*	*92,558*
• Compression Unit			
CO_2 Compressors	kW_e	48,790	62,165
Sub Total	*kW_e*	*48,790*	*62,165*
Net Power Output	**kW_e**	**549,960**	**549,969**
• Net Plant Efficiency (HHV)	%	26.2	23.1
• Net Plant Heat Rate	Btu/kWh	13,046	14,785
Condenser Cooling Duty	**MMBTU/hr**	**1,928**	**1,716**
Consumables			
• As-Received Coal Feed	lb/hr	614,994	697,032
• Thermal Input[4]	kW_{th}	2,102,643	2,383,129
Flue Gas Conditions before Carbon Capture Facility			
• Flow Rate	lb/hr	7,084,027	8,014,057
• Temperature	°F	136	136
• Composition			
Ar	% mole	0.0081	0.0081
CO_2	% mole	0.1350	0.1350
H_2	% mole	0.0000	0.0000
H_2O	% mole	0.1537	0.1537
N_2	% mole	0.6793	0.6793
O_2	% mole	0.0238	0.0238
SO_2	% mole	0.0000	0.0000
Total	% mole	1.0000	1.0000

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[3] No breakdown is given for the total auxiliary load of the amine-based carbon capture facility.
[4] HHV of As-Received Illinois No.6 coal is 11,666 Btu/lb.

Parameter	Unit	Amine Technology	Solid Sorbent Technology
CO_2 Capture Profile			
• CO_2 Produced	lb/hr	1,459,838	1,651,831
• CO_2 Captured	lb/hr	1,313,960	1,486,648
• CO_2 Emission	lb/hr	145,878	165,183
Total Circulating Cooling Water Demand	**gpm**	**540,000**	**487,120**
• PC Plant	gpm	200,000	226,679
• Carbon Capture Facility	gpm	340,000	260,441
Steam Extraction from IP/LP Turbine before Carbon Capture Facility			
• Flow Rate	lb/hr	1,931,497	2,709,721
• Temperature	°F	565	563
• Pressure	psia	73.5	45.0
• Total Regeneration Energy Requirement	MMBtu/hr	2,010	3,036
• Thermal Regeneration Energy	Btu/lb CO_2	1,530	2,042
• Amount of Amine Solvent/Solid Sorbent	lb/hr	Not Given	27,530,510

3.3 MAJOR EQUIPMENT LIST

Tables below are major equipment items for the subcritical pulverized coal-fired power plant integrated with the solid sorbent-based carbon capture facility.

Table 3-2 Account 1: Fuel and Sorbent Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Bottom Trestle Dumper and Receiving Hoppers	N/A	205 tonne (226 ton)	2	0
2	Feeder	Belt	648 tonne (715 ton)	2	0
3	Conveyor No. 1	Belt	1285 tonne/hr (1417 tph)	1	0
4	Transfer Tower No. 1	Enclosed	N/A	1	0
5	Conveyor No. 2	Belt	1285 tonne/hr (1417 tph)	1	0
6	As-Received Coal Sampling System	Two-stage	N/A	1	0
7	Stacker / Reclaimer	Traveling, linear	1285 tonne/hr (1417 tph)	1	0
8	Reclaim Hopper	N/A	61 tonne (67 ton)	2	1
9	Feeder	Vibratory	257 tonne/hr (284 tph)	2	1
10	Conveyor No. 3	Belt w/ tripper	525 tonne/hr (578 tph)	1	0
11	Crusher Tower	N/A	N/A	1	0
12	Coal Surge Bin w/ Vent Filter	Dual outlet	257 tonne/hr (284 tph)	2	0
13	Crusher	Impactor reduction	8.5 cm x 0 - 3.2 cm x 0 (3.4 in x 0 -1.3 in x 0)	2	0
14	As-Fired Coal Sampling System	Swing hammer	N/A	1	1
15	Conveyor No. 4	Belt w/ tripper	525 tonne/hr (578 tph)	1	0
16	Transfer Tower No. 2	Enclosed	N/A	1	0
17	Conveyor No. 5	Belt w/ tripper	525 tonne/hr (578 tph)	1	0
18	Coal Silo w/ Vent Filter and Slide Gates	Field erected	1131 tonne/hr (1247 tph)	3	0
19	Limestone Truck Unloading Hopper	N/A	41 tonne/hr (45 tph)	1	0
20	Limestone Feeder	Belt	134 tonne/hr (147 tph)	1	0
21	Limestone Conveyor No. L1	Belt	134 tonne/hr (147 tph)	1	0
22	Limestone Reclaim Hopper	N/A	31 tonne/hr (34 tph)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
23	Limestone Reclaim Feeder	Belt	103 tonne/hr (114 tph)	1	0
24	Limestone Conveyor No. L2	Belt	103 tonne/hr (114 tph)	1	0
25	Limestone Day Bin	w/ actuator	422 tonne/hr (465 tph)	2	0

Table 3-3 Account 2: Fuel and Sorbent Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Coal Feeder	Gravimetric	61 tonne/hr (67 tph)	6	0
2	Coal Pulverizer	Ball type or equivalent	61 tonne/hr (67 tph)	6	0
3	Limestone Weigh Feeder	Gravimetric	35 tonne/hr (39 tph)	1	1
4	Limestone Ball Mill	Rotary	35 tonne/hr (39 tph)	1	1
5	Limestone Mill Slurry Tank with Agitator	N/A	137300 liters (36300 gal)	1	1
6	Limestone Mill Recycle Pumps	Horizontal centrifugal	2274 lpm @ 12m H2O (601 gpm @ 40 ft H2O)	1	1
7	Hydroclone Classifier	4 active cyclones in a 5 cyclone bank	558 lpm (147 gpm) per cyclone	1	1
8	Distribution Box	2-way	N/A	1	1
9	Limestone Slurry Storage Tank with Agitator	Field erected	763686 liters (202000 gal)	1	1
10	Limestone Slurry Feed Pumps	Horizontal centrifugal	1588 lpm @ 9m H2O (419 gpm @ 30 ft H2O)	1	1

Table 3-4 Account 3: Feedwater and Miscellaneous Systems and Equipment

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Demineralized Water Storage Tank	Vertical, cylindrical, outdoor	1771700 liters (468000 gal)	2	0
2	Condensate Pumps	Vertical canned	18145 lpm @ 213m H2O (4794 gpm @ 700 ft H2O)	1	1
3	Deaerator and Storage Tank	Horizontal spray type	2946886 kg/hr (6496764 lb/h), 5 min. tank	1	0
4	Boiler Feed Pump / Turbine	Barrel type, multi-stage, centrifugal	49331 lpm @ 2591 m H2O (13032 gpm @ 8500 ft H2O)	1	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
5	Startup Boiler Feed Pump, Electric Motor Driven	Barrel type, multi-stage, centrifugal	14585 lpm @ 2591 m H2O (3853 gpm @ 8500 ft H2O)	1	0
6	LP Feedwater Heater 1A/1B	Horizontal U-tube	543600 kg/hr (1198400 lb/hr)	2	0
7	LP Feedwater Heater 2A/2B	Horizontal U-tube	543600 kg/hr (1198400 lb/hr)	2	0
8	LP Feedwater Heater 3A/3B	Horizontal U-tube	543600 kg/hr (1198400 lb/hr)	2	0
9	LP Feedwater Heater 4A/4B	Horizontal U-tube	543600 kg/hr (1198400 lb/hr)	2	0
10	HP Feedwater Heater 6	Horizontal U-tube	2945300 kg/hr (6493400 lb/hr)	1	0
11	HP Feedwater Heater 7	Horizontal U-tube	2945300 kg/hr (6493400 lb/hr)	1	0
12	Auxiliary Boiler	Shop fabricated, water tube	18144 kg/hr, 2.8 Mpa, 343 C (40000 lb/hr, 400 psig, 650 F)	1	0
13	Fuel Oil System	No. 2 fuel oil for light off	1135624 litre (300000 gal)	1	0
14	Service Air Compressors	Flooded screw	28 m3/min @ 0.7 Mpa (1000 scfm @ 100 psig)	2	1
15	Instrument Air Dryers	Duplex, regenerative	28 m3/min (1000 scfm)	2	1
16	Closed Cycle Cooling Heat Exchangers	Shell and tube	53 GJ/hr (50 MMBtu/hr) each	2	0
17	Closed Cycle Cooling Water Pumps	Horizontal centrifugal	20820 lpm @ 30 m H2O (5500 gpm @ 100 ft H2O)	2	1
18	Engine-Driven Fire Pump	Vertical turbine, diesel engine	3785 lpm @ 88 m H2O (1000 gpm @ 290 ft H2O)	1	1
19	Fire Service Booster Pump	Two-stage horizontal centrifugal	2650 lpm @ 64 m H2O (700 gpm @ 210 ft H2O)	1	1
20	Raw Water Pumps	Stainless steel, single suction	9410 lpm @ 18 m H2O (2486 gpm @ 60 ft H2O)	2	1
21	Ground Water Pumps	Stainless steel, single suction	3768 lpm @ 268 m H2O (995 gpm @ 880 ft H2O)	5	1
22	Filtered Water Pumps	Stainless steel, single suction	2480 lpm @ 49 m H2O (655 gpm @ 160 ft H2O)	2	1
23	Filtered Water Tank	Vertical, cylindrical	2379648 litre (628636 gal)	1	0
24	Markup Water Demineralizer	Multi-media filter, cartride filter, RO membrane assembly, electrodeionization unit	814 lpm (215 gpm)	1	1

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
25	Liquid Waste Treatment System	-	10 years, 24-hour storm	1	0

Table 3-5 Account 4: Boiler and Accessories

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Boiler	Subcritical, drum wall-fired, low Nox burners, overfire air	2945342 kg/hr steam @ 17.9 Mpa/574 C/574 C (6493000 lb/hr steam @ 2600 psig/1065 F/1065 F)	1	0
2	Primary Air Fan	Centrifugal	401514 kg/hr, 5482 m3/min @ 123 cm WG (885000 lb/hr, 194000 acfm @ 48 in. WG)	2	0
3	Forced Draft Fan	Centrifugal	1307353 kg/hr, 17844 m3/min @ 47 cm WG (2882000 lb/hr, 630000 acfm @ 19 in. WG)	2	0
4	Induced Draft Fan	Centrifugal	1893352 kg/hr, 39958 m3/min @ 104 cm WG (4174000 lb/hr, 1411000 acfm @ 41 in. WG)	2	0
5	SCR Reactor Vessel	Space for spare layer	3794100 kg/hr (8364500 lb/hr)	2	0
6	SCR Catalyst	-	-	3	0
7	Dilution Air Blower	Centrifugal	224 m3/min @ 108 cm WG (7900 acfm @ 42 in. WG)	2	1
8	Ammonia Storage	Horizontal tank	248800 liters (65700 gal)	5	0
9	Ammonia Feed Pump	Centrifugal	48 lpm @ 91 m H2O (13 gpm @ 300 ft H2O)	2	1

Table 3-6 Account 5: Flue Gas Cleanup

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Fabric Filter	Single stage, high ratio with pulse jet online cleaning system	1896500 kg/hr (4181100 lb/hr), 99.8% efficiency	2	0
2	Absorber Module	Counter current open	75800 m3/min (2678000 acfm)	1	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
		spray			
3	Recirculation Pumps	Horizontal centrifugal	261710 lpm @ 64 m H2O (69100 gpm @ 210 ft H2O)	5	1
4	Bleed Pumps	Horizontal centrifugal	6910 lpm (1820 gpm) at 20 wt% solids	2	1
5	Oxidation Air Blowers	Centrifugal	133 m3/min @ 0.3 MPa (4680 acfm @ 37 psia)	2	1
6	Agitators	Side entering	50 hp	5	1
7	Dewatering Cyclones	Radial assembly, 5 units each	1716 lpm (450 gpm) per cyclone	2	0
8	Vacuum Filter Belt	Horizontal belt	54 tonne/hr (60 tph) of 50 wt% slurry	2	1
9	Filtrate Water Return Pumps	Horizontal centrifugal	1029 lpm @ 12 m H2O (270 gpm @ 40 ft H2O)	1	1
10	Filtrate Water Return Storage Tank	Vertical, lined	686460 lpm (181000 gal)	1	0
11	Process Markup Water Pumps	Horizontal centrifugal	5534 lpm @ 21 m H2O (1460 gpm @ 70 ft H2O)	1	1

Table 3-7 Account 5B: Carbon Dioxide Recovery

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	ADA-ES Capture Process	Solid Absorbent	90878 kg/hr (200351 lb/hr), 20.6 wt % CO2 concentration	44 trains	0
2	Condensate Pump back to Power Island	Centrifugal	25860 lpm @ 52 m H2O (6800 gpm @ 170 ft H2O)	1	1
3	CO_2 Compressor	Integrally geared, multi-stages centrifugal	199621 kg/hr @ 15.3 MPa (440088 lb/hr @ 2215 psia)	4	0
4	Dehydration	Desiccant (Activated Alumina)	199621 kg/hr @ 2.75 MPa (440088 lb/hr @ 400 psia)	4	0

Table 3-8 Account 7: HRSG, Ducting & Stack

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Stack	Reinforced concrete with FRP liner	152 m (500 ft) high x 6 m (20 ft) diameter)	1	0

Table 3-9 Account 8: Steam Turbine Generator and Auxiliaries

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Steam Turbine (Steam Turbine in Power Island + Backpressure Turbine at Upstream of ADA-ES Reboiler)	Commercially available advanced steam turbine	756 MW, 16.5 MPa/566 C/566 C (2400.3 psig/1050 F/1050 F) Plus 34 MW, 0.52 MPa/295 C (75 psia to 45 psia/565 F)	1	0
2	Steam Turbine Generator	Hydrogen cooled, static excitation	882 MVA @ 0.9 p.f., 24 kV, 60 Hz, 3 phase	1	0
3	Surface Condenser	Single pass, divided waterbox including vacuum pumps	1986 GJ/hr (1882 MMBtu/hr), Inlet water temperature 16 C (60 F), Water temperature rise 11 C (20 F)	1	0

Table 3-10 Account 9: Cooling Water System

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Circulating Water Pumps	Vertical, wet pit	827488 lpm @ 30 m H2O (218599 gpm @ 100 ft H2O)	2	1
2	Cooling Tower	Evaporative, mechanical draft, multi cell	11 C (51.5 F) wet bulb / 16 C (60 F) CWT / 27 C (80 F) HWT / 4623 GJ/hr (4382 MMBtu/hr) heat duty	1	0

Table 3-11 Account 10: Ash/Spent Sorbent Recovery and Handling

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	Economizer Hopper (part of boiler scope of supply)	-	-	4	0
2	Bottom Ash Hopper (part of boiler scope of supply)	-	-	2	0
3	Clinker Grinder	-	7.3 tonne/hr (8 tph)	1	1
4	Pyrites Hopper (part of pulverizer scope of supply included with boiler)	-	-	6	0
5	Hydroejectors	-	-	12	0

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
6	Economizer / Pyrites Transfer Tank	-	-	1	0
7	Ash Sluice Pumps	Vertical, wet pit	257 lpm @ 17 m H2O (68 gpm @ 56 ft H2O)	1	1
8	Ash Seal Water Pumps	Vertical, wet pit	8581 lpm @ 9 m H2O (2270 gpm @ 28 ft H2O)	1	1
9	Hydrobins	-	257 lpm (68 gpm)	1	1
10	Baghouse Hopper (part of baghouse scope of supply)	-	-	24	0
11	Air Heater Hopper (part of boiler scope of supply)	-	-	10	0
12	Air Blower	-	25 m3/min @ 0.2 MPa (873 scfm @ 24 psi)	1	1
13	Fly Ash Silo	Reinforced concrete	1645 tonne (1813 ton)	2	0
14	Slide Gate Valve	-	-	2	0
15	Unloader	-	-	1	0
16	Telescoping Unloading Chute	-	154 tonne/hr (170 tph)	1	0

Table 3-12 Account 11: Accessory Electric Plant

Equipment No.	Description	Type	Design Condition	Operating Qty.	Spares
1	STG Transformer	Oil-filled	24 kV/345 kV, 726 MVA, 3-ph, 60 Hz)	1	0
2	Auxiliary Transformer	Oil-filled	24 kV/4.16 kV, 150 MVA, 3-ph, 60 Hz)	1	1
3	Low Voltage Transformer	Dry ventilated	4.16 kV/480 V, 22 MVA, 3-ph, 60 Hz)	1	1
4	STG Isolated Phase Bus Duct and Tap Bus	Aluminum, self-cooled	24 kV, 3-ph, 60 Hz	1	0
5	Medium Voltage Switchgear	Metal clad	4.16 kV, 3-ph, 60 Hz	1	1
6	Low Voltage Switchgear	Metal enclosed	480 V, 3-ph, 60 Hz	1	1
7	Emergency Diesel Generator	Sized for emergency shutdown	750 kW, 480 V, 3-ph, 60 Hz	1	0

Table 3-13 Account 12: Instrumentation and Control

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

3.4 EQUIPMENT USEFUL LIFE

Generally, the design of equipment for post combustion treatment of flue gases targets a useful life of 30 years. Material selection is chosen to adequately provide this life span, but in practice, longer lifetimes would be expected

Pretreatment of the flue gas based on flue gas desulphurization has a long operative history in utilities, and a life span of 30 years should be readily achievable. Assuming a reasonable amount of chlorides, the design of the wet / dry interface materials will likely feature a great deal of Alloy 2205 for pitting resistance. Recirculation piping would consist of erosion resistant fiberglass or reasonable substitute. Mechanical devices would have normal maintenance requirements but generally should last the expected life of the facility.

The sorbent absorber, regenerator, and interconnected risers are constructed of carbon steel. Areas of high erosive potential, namely the cyclone separators on the absorber and regenerator, feature an erosion resistant refractory lining. Without the benefit of long term testing, this selection appears to be satisfactory. At full scale, areas of concern might include parts of the risers where direction is changed at elbows.

The flue gas compressor used to pressurize the feed flue gas would also be specified to achieve a 30 year life. A major overhaul can be expected every 6 to 7 years.

Recycle compressors used for internal recirculation of carbon dioxide are expected to be similar to the flue gas compressor, just on a much smaller scale, provided they are kept free of particulate.

The carbon dioxide compressor would also mirror the life expectancy of the flue gas compressor.

4.0 ECONOMIC ANALYSIS

The methodology for capital costs, operating and maintenance costs and the first year Cost of Electricity (COE) used in this report are based on the DOE/NETL Final Report[1]. Background information directly relevant to the analysis is briefly given in each section. All the costs are estimated in October 2015 US dollars. Appendix C provides the background calculations for this methodology.

4.1 CAPITAL COST ESTIMATE

Capital costs of the PC plant are obtained by linearly scaling the costs of Case A (escalated from 2007 to 2015 US dollars) proportional to the coal flow rate. Total inflation over the period from June 2007 to October 2015 is 14.15 percent. Final results are also presented as February 2011 dollars for comparison to earlier developed costs.

The cost of the process equipment for the carbon capture facility, including the SO_2 polishing unit, the BP turbine and the CO_2 compressor and dehydration, are estimated using a combination of vendor quotes and actual pricing information from Stantec's internal database.

As determined by the Association for the Advancement of Cost Engineering International (AACE), the level of cost accuracy for this concept design is defined by a Class IV/V estimate, which is approximately in a range of ±50 percent.

Items specifically excluded from this capital cost estimate are the same as those in Case A – taxes, site specific considerations, overtime and extra premiums due to an Engineering, Procurement, and Construction (EPC) type contract.

Four levels of capital cost and the estimate basis used in this analysis are given in Table 4-1. Each "level" flows through to the next. From the table, the BEC is the cost of process equipment plus other costs (e.g. cost of supporting facilities and direct/indirect labor costs related to construction and installation). The core equipment costs for Cases A and Cases B are identical. The method of calculating indirects for the final BEC cost of the carbon capture facility utilizes the DOE method to build up costs and includes indirects[1].

Table 4-1 Capital Cost Structure

Level	Cost	Estimate Basis
Bare Erected Cost (BEC) – Overnight cost expressed in base year (2011) dollars	- Process Equipment	
	- Support Facilities (e.g. Offices, Labs)	
	- Direct and Indirect Labor for Construction and/or Installation	
Total Plant Cost (TPC) – Overnight cost expressed in base year (2011) dollars	- BEC	
	- Engineering, Procurement and Construction (EPC) Contractor Services	~ 9.4% of BEC
	- Process Contingency	~ 20% of CO_2 Capture Facility Capital Cost
	- Project Contingency	~ 15% of the Sum of BEC, EPC and Process Contingency
Total Overnight Cost (TOC) – Overnight cost expressed in base year (2011) dollars	- TPC	
	- Preproduction	~ 3% of TPC
	- Inventory Capital	~ 2% of TPC
	- Initial Cost	
	- Land	
	- Owner's Costs	15% of TPC
	- Financing Costs	2.7% of TPC
Total As-Spent Cost (TOC) – Expressed in mixed-year current dollars over the capital expenditure period	- TOC	
	- Escalation/Interest During Capital Expenditure Period	1.140 Multiplier for High Risk Investor-Owned Utility (IOU) Over Five-Year Capital Expenditure Period

Table 4-2 and 4-3 summarize the capital cost estimate results. Appendix C contains the complete capital cost table for the carbon capture facility for both cases. Details on Case A will be discussed in Section 5.0.

Table 4-2 Capital Cost Summary 2015 Dollars

Description	Unit	CASE B
1. PC Plant	US$	1,136,090,912
2. Carbon Capture Facility[*]	US$	2,139,215,668
3. CO_2 Compression and Dehydration	US$	102,688,141
BEC	**US$**	**3,377,994,721**
4. ECM	US$	318,845,653
5. Process Contingency	US$	435,362,068
6. Project Contingency	US$	592,750,828
TPC	**US$**	**4,724,953,270**
7. Preproduction Costs	US$	151,119,850
8. Inventory Capital	US$	96,819,840
9. Initial Cost for Catalyst and Chemicals[†]	US$	180,802,311
10. Land[ξ]	US$	1,027,368
11. Other Owner's Costs	US$	708,743,255
12. Financing Costs	US$	127,575,374
TOC	**US$**	**5,991,041,269**
TASC Multiplier (IOU, 35 Years)	-	1.140
TASC	**US$**	**6,355,138,125**

The cost data presented as 2011 Dollars is:

Table 4-3 Capital Cost Summary 2011 Dollars

Description	Unit	CASE B
1. PC Plant	US$	1,057,136,133
2. Carbon Capture Facility*	US$	1,990,546,844
3. CO_2 Compression and Dehydration	US$	95,551,636
BEC	**US$**	**3,143,234,612**

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[*] Included cost of the SO_2 polishing unit and the BP turbine.

[†] Summarized in Table 4-3 for both cases and see more details in Appendix B.

[ξ] Assumed the same cost of land as Case A.

4. ECM	US$	296,686,873.52
5. Process Contingency	US$	405,105,760.51
6. Project Contingency	US$	551,556,492
TPC	**US$**	**4,396,583,739**
7. Preproduction Costs	US$	140,617,491
8. Inventory Capital	US$	90,091,163
9. Initial Cost for Catalyst and Chemicals†	US$	168,237,113
10. Land[ξ]	US$	955,969
11. Other Owner's Costs	US$	659,487,807
12. Financing Costs	US$	118,709,283
TOC	**US$**	5,574,682,566
TASC Multiplier (IOU, 35 Years)	-	1.14
TASC	**US$**	**6,355,138,125**

4.2 OPERATING AND MAINTENANCE (O&M) COST ESTIMATE

O&M costs for Case B, summarized in Table 4-4 and 4-5, are developed based on the following:

Fixed O&M Costs

- Annual operating labor cost / administrative and support labor are equal to the escalated costs of Case A.
- Maintenance labor cost / property taxes and insurance are obtained by linearly scaling the escalated costs of Case A proportional to the TPC.

Variable O&M Costs

- Annual costs are calculated at 85 percent capacity factor.
- Water consumption is estimated using the Case A calculation method.
- Maintenance material cost / annual cost of MU and WT chemicals are obtained by linearly scaling the escalated costs of Case A proportional to the TPC.
- Annual costs of limestone / ammonia / SCR catalyst / fly ash and bottom ash disposal are obtained by linearly scaling the escalated costs of Case A proportional to the gross output.

[ξ] Assumed the same cost of land as Case A.

- Initial and annual costs of sodium hydroxide (NaOH) / sulfuric acid (H_2SO_4) are obtained by linearly scaling the escalated costs of Case A proportional to the coal flow rate. The SO_2 concentration must be less than 1 ppm for the solid sorbent-based process, resulting in a higher NaOH consumption than the amine process. Therefore, annual NaOH cost is doubled to reflect the high consumption.

- Consumption of NaOH / Lime / H_2SO_4 for additional waste water treatment depends on the cooling water demand of the carbon capture facility.

- Makeup sorbent is required to replenish loss due to attrition. Solid sorbent loss of 0.005 percent per cycle (attrition) is assumed. A leakage loss is not accounted for.

- The cost of solid sorbents is US$5 per pound.

- Consumable pricing (caustic, lime, sulphuric acid, etc.) can be found in Appendix C. Costs extracted from the DOE report[1] for consumables are adjusted to October 2015 dollars with a total inflation rate from June 2007 to October 2015 of 14.15 percent.

- Potential revenue from selling CO_2 product is excluded.

Table 4-4 Operating Cost Summary 2015 Dollars

Parameters	Case B	
	Initial Cost (US$)	Annual Cost (US$/year)
Fixed O&M Costs	N/A	141,311,446
Variable O&M Costs	180,802,311	132,049,027
Fuel Cost	0	113,110,975
Total	**180,802,311**	**386,471,468**

Table 4-5 Operating Cost Summary 2011 Dollars

Parameters	Case B	
	Initial Cost (US$)	Annual Cost (US$/year)
Fixed O&M Costs	N/A	131,490,741
Variable O&M Costs	168,237,113	125,219,297
Fuel Cost	0	105,250,115
Total	**168,237,113**	**361,960,153**

4.3 CO_2 TRANSPORT, STORAGE AND MONITORING (TS&M) COST

Supercritical CO_2 from the sorbent based carbon capture facility is transported via pipeline and injected to a deep saline formation for sequestration. Since a set of assumptions used in modeling the TS&M cost for Case A is applied to Case B, the CO_2 TS&M costs for Cases B linearly depend on the CO_2 removal rate. Results show approximately US$30 M for both cases.

4.4 COST OF ELECTRICITY (COE) CALCULATION

4.4.1 Methodology

In this report, the methodology developed by DOE / NETL is used to calculate a COE and a Levelized COE (LCOE)[1]. Both COE and LCOE are the revenue received by the generator per net megawatt-hour (MWh) during the PC plant’s first year of operation. The calculated COE and LCOE are expected to provide an Internal Rate of Return on Equity (IRROE) when the calculation is conducted over the entire economic analysis period as defined in Table 4-6. The difference is that the COE escalates thereafter at a nominal annual rate equal to the general inflation rate of three percent, whereas the LCOE escalates thereafter at a zero nominal annual rate.

Table 4-7 provides details on finance structures of the PC plant with carbon capture, assuming high risk Investor-Owned Utility (IOU).

Table 4-6 Global Economic Assumption

Parameter	Value
Taxes	
Income Tax Rate	38% (Effective 34% Federal, 6% State)
Capital Depreciation	20 years, 150% declining balance
Investment Tax Credit	0%
Tax Holiday	0 years
Contracting and Financing Terms	
Contracting Strategy	Engineering, Procurement, Construction and Management (EPCM)
Type of Debt Financing	Non-Recourse
Repayment Term of Debt	15 years
Grace Period on Debt Repayment	0 years
Debt Reserve Fund	None
Analysis Time Periods	
Capital Expenditure Period	5 years
Operation Period	30 years
Economic Analysis Period	35 years
Treatment of Capital Costs	
Capital Cost Escalation During Capital Expenditure Period (nominal annual rate)	3.6%
Distribution of TOC Over Capital Expenditure Period (before escalation)	5-year period: 10%, 30%, 25%, 20%, 15%
Working Capital	Zero for all parameters
% of TOC that is Depreciated	100%
Escalation of Operating Revenues and Costs	
Escalation of COE (revenue), O&M Costs and Fuel Costs (nominal annual rate)	3.0%

Table 4-7 Financial Structure for High Rish IOU

Parameters	%	Current (Nominal) Dollar Cost	Weighted Current (Nominal) Cost	After Tax Weighted Cost of Capital
Debt	45	5.5%	2.475%	
Equity	55	12%	6.6%	
Total	**100**		**9.075%**	**8.13%**

4.4.2 Cost of Electricity Formulas

The COE is calculated by the following equation published in the DOE report[1]. All costs used in this equation are expressed in base-year dollars. The base year is assumed to be 2011, which is the first year of capital expenditure. Since the COE and O&M and fuel costs escalate thereafter at three percent nominal annual general inflation rate, all costs expressed in first-year-of-operation dollars are equivalent to base-year costs. As a result, the first-year COE is also equivalent to the base year COE.

$$\mathrm{COE} = \frac{(\mathrm{CCF})(\mathrm{TOC}) + (\mathrm{OC}_{\mathrm{FIX}}) + (\mathrm{CF})(\mathrm{OC}_{\mathrm{VAR}})}{(\mathrm{CF})(\mathrm{MWh})} \qquad \text{(Eq.1)}$$

where

COE = revenue received by the generator per net megawatt-hour during the PC plant's first year of operation (but expressed in base-year dollars, \$/MWh or equivalent to mills/kWh).

CCF = capital charge factor, which was 0.124 for high risk IOU, 5-year capital expenditure period.

TOC = total overnight cost, expressed in base-year dollars.

OCFIX = the sum of all fixed annual operating costs, expressed in base year dollars.

OCVAR= the sum of all variable annual operating costs, including fuel at 100% capacity factor, expressed in base year dollars.

CF = plant capacity factor, assumed to be constant over the operational period.

MWh = annual net megawatt-hours of power generated at 100% capacity factor.

4.4.3 Results

Both base year and first operation year COE / LCOE for Case B are summarized in Table 4-8 and 4-9.

Table 4-8 COE and LCOE Summary in 2015 Dollars

Parameter	Case B
COE (mills/kWh, base year 2015$)	283.9
Capital costs	181.4
Fixed O&M Costs	34.5
Variable O&M Costs	32.9
Fuel Costs	27.6
CO_2 T&SM Costs	7.5
LCOE (mills/kWh, base year 2015$)	359.9
COE (mills/kWh, first operation year 2020$)	329.1
LCOE (mills/kWh, first operation year 2020$)	417.2

Table 4-9 COE and LCOE Summary in 2011 Dollars

Parameter	Case B
COE (mills/kWh, base year 2011$)	264.2
Capital costs	168.8
Fixed O&M Costs	32.1
Variable O&M Costs	30.6
Fuel Costs	25.7
CO_2 T&SM Costs	7.0
LCOE (mills/kWh, base year 2011$)	334.9
COE (mills/kWh, first operation year 2016$)	306.2
LCOE (mills/kWh, first operation year 2016$)	388.2

5.0 COMPARABLE CARBON CAPTURE TECHNOLOGIES

In this section, the developed costs for Case B from Section 4 are compared to the established costs for Case A from the DOE report[1].

Table 5-1 Performance and Economic Comparison in 2015 Dollars*

Description	Unit	Case A	Case B
PERFORMANCE			
1. Gross Power Output	kW_e	672,700	718,738
2. Electrical Output – BP Turbine	kW_e	0	32,525
3. Auxiliary Load Summary	kW_e	122,740	
• PC Plant	kW_e	51,550	43,722
• SO_2 Polishing Unit	kW_e	0	2,845
• Carbon Capture Facility	kW_e	22,400	92,558
• Compression / Dehydration Unit	kW_e	48,790	62,165
4. Net Power Output	kW_e	549,960	549,969
5. Thermal Input	kW_{th}	2,102,644	2,383,129
• Gross Output – PC Plant	kW_{th}	1,513,467	-
• Sorbent Regeneration	kW_{th}	589,177	-
6. Coal Flow Rate	lb/hr	614,994	697,032
7. Coal HHV	Btu/lb	11,666	11,666
8. Net Plant HHV Efficiency	%	26.2	23.1
9. Net Plant HHV Heat Rate	Btu/kWh	13,046	14,785
10. Net kWh @ 85% Capacity Factor	kWh-net	4,095,002,160	4,095,075,837
11. CO_2 Capture Profile			
• CO_2 Produced	lb/hr	1,459,838	1,651,831
• CO_2 Captured	lb/hr	1,313,960	1,486,648
• CO_2 Emission	lb/hr	145,878	165,183
12. Circulating Cooling Water Demand[†]	gpm	540,000	487,120
• CW Demand – PC Plant	gpm	200,000	226,679
• CW Demand – Capture Facility	gpm	340,000	260,441
13. Raw Water Consumption	1000 gal/D	8,081	7,802
14. Carbon Capture Facility			
• Solvent Regeneration Energy	Btu/lb CO_2	1,530	2,042
• Total Energy Requirement	MMBtu/hr	2,010	3,036
• Steam Extraction Flow	lb/hr	1,931,497	2,709,721
• Amount of Solid Sorbent	lb/hr	N/A	27,530,510

[†] Assumed maximum circulating CW demands required by the PC plant and the carbon facility for Case A.

Description	**Unit**	**Case A**		**Case B**	
COSTS					
A. Capital Cost					
1. Bare Erected Cost	$	1,415,830,829		3,377,994,721	
• PC Plant	$	1,047,285,351		1,136,090,912	
• Carbon Capture Facility	$	325,099,223		2,139,215,668	
• CO_2 Compression / Dehydration	$	43,446,256		102,688,141	
2. Total Plant Cost	2015$	1,847,389,360		4,724,953,270	
3. Total Plant Cost	2015$/kW	3,359		8,591	
4. Total Overnight Cost	2015$	2,266,413,999		5,991,041,269	
5. Total Overnight Cost	2015$/kW	4,121		10,893	
6. Total As-Spent Cost	2015$	2,583,711,959		6,829,787,046	
7. Total As-Spent Cost	2015$/kW	4,698		12,417	
B. Operating & Maintenance Cost		**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**
1. Total Fixed Operating Costs	$	N/A	61,025,905	N/A	141,311,446
2. Total Variable Operating Costs	**$**	**3,096,311**	**42,801,959**	**180,802,311**	**134,571,604**
• Maintenance Material Cost	$	-	**17,858,299**	-	**47,206,648**
• Consumables	$	3,096,311	**24,943,660**	180,802,311	**87,364,956**
C. Fuel Costs	**$**	**99,798,295**		**113,110,975**	
D. CO_2 TS&M Costs	$	**27,112,265**		**30,675,509**	
E. COE contribution (2015$)	**mills/kWh**	**125.0**		**283.9**	
• Capital Costs	mills/kWh	68.6		181.4	
• Fixed Operating Costs	mills/kWh	14.9		34.5	
• Variable Operating Costs	mills/kWh	10.5		32.9	
• Fuel Costs	mills/kWh	24.4		27.6	
• CO_2 T&SM Costs	mills/kWh	6.6		7.5	
F. LCOE (2015$)	**mills/kWh**	**158.7**		**359.9**	

Table 5-2 Economic Comparison in 2011 Dollars

Description	**Unit**	**Case A**		**Case B**	
COSTS					
A. Capital Cost					
1. Bare Erected Cost	$	1,317,434,997		3,143,235,000	
• PC Plant	$	974,502,282		1,057,136,133	
• Carbon Capture Facility	$	302,505,840		1,990,546,844	
• CO$_2$ Compression / Dehydration	$	40,426,877		95,551,636	
2. Total Plant Cost	2011$	1,719,001,555		4,396,584,000	
3. Total Plant Cost	2011$/kW	3,126		7,994	
4. Total Overnight Cost	2011$	2,108,905,287		5,574,683,000	
5. Total Overnight Cost	2011$/kW	3,835		10,136	
6. Total As-Spent Cost	2011$	2,404,152,028		6,355,138,000	
7. Total As-Spent Cost	2011$/kW	4,372		11,554	
B. Operating & Maintenance Cost		**Initial Cost**	**Annual Cost**	**Initial Cost**	**Annual Cost**
1. Total Fixed Operating Costs	$	N/A	56,784,795	N/A	131,490,741
2. Total Variable Operating Costs	**$**	**2,881,127**	**39,827,356**	**168,237,114**	**125,219,297**
• Maintenance Material Cost	$	-	**16,617,203**	-	**43,925,933**
• Consumables	$	2,881,127	**23,210,153**	168,237,114	**81,293,364**
C. Fuel Costs	**$**	**92,862,624**		**105,250,115**	
D. CO$_2$ TS&M Costs	$	**25,228,047**		**28,543,657**	
E. COE contribution (2011$)	**mills/kWh**	**116.3**		**264.2**	
• Capital Costs	mills/kWh	63.8		168.8	
• Fixed Operating Costs	mills/kWh	13.9		32.1	
• Variable Operating Costs	mills/kWh	9.8		30.6	
• Fuel Costs	mills/kWh	22.7		25.7	
• CO$_2$ T&SM Costs	mills/kWh	6.1		7.0	
F. LCOE (2011$)	**mills/kWh**	**147.7**		**334.9**	

The cost per ton of carbon dioxide captured is presented below in Table 5-3.

Table 5-3 $/ton CO_2 Captured Results

	Case A	Case B
2015 Dollars	133 $/ton	256 $/ton
2011 Dollars	124 $/ton	239 $/ton

6.0 CONCLUSIONS

As can be seen in previous sections, the solid sorbent based capture technology does not compare favorably to conventional amine based absorption. The LCOE, a factor comparing both capital and operating costs, is larger for solid sorbents by a factor of 2.2. There a many contributing factors to this higher cost of energy, but primarily the central issue is the lower working capacity discovered during the pilot.

On the capital side, the pilot results lead to a significant increase in the number of parallel treatment trains. The original TEA was based on a higher working capacity, which only required 4 parallel absorber / desorber trains. The actual residence time developed by the pilot of 65 minutes resulted in a substantial increase in mass flow of sorbent. Fluidizing these solids limits how deep the beds can be, so additional contact can only be provided by adding more parallel trains. Based on pilot results, 44 trains are necessary for 65 minutes of residence time.

Eliminating the latent heat portions of regenerating a solid sorbent is the promising advantage of the technology over amine based, water solution absorption. The growing volume of sorbent necessary for 65 minutes of contact time results in a significant increase in the amount of steam for regeneration. This study pins the output of the power plant at 550 MW net at all times, so as steam for regeneration increases, the gross output of the plant must also increase, leading to higher capital and operating costs for the base power plant.

On the operating side, due to attrition losses being calculated on the total mass flow of sorbent, the higher solids flow rate does result in a higher replacement cost of sorbent. Also attributed to the growth of the power plant, fuel costs rise, again leading to a higher LCOE for the solid sorbent case.

Based on this assessment, solid sorbents are not competitive with amine based technology at this time. In order to restore competitiveness, the working capacity of the sorbent must be increased.

7.0 REFERENCES

1 Cost and Performance Baseline for Fossil Energy Plants – Volume 1: Bituminous Coal and Natural Gas to Electricity, Revision 3, July, 6, 2015, DOE/NETL-2015/1723.

2 Techno-Economic Report Commercial-Scale Solid Sorbent CO_2 Capture Process: 550 MW Pulverized Coal-Fired Power Plant, July 12, 2011, Stantec.

8.0 APPENDICES

Appendix A PILOT PLANT PROCESS FLOW DIAGRAM

Appendix B GATECYCLE RESULTS

Table B-1 Comparable GateCycle Results with DOE's Cases 9 and 10

	Bituminous Baseline Study (DOE/NETL, 2010)		**Process Validation**	
	Subcritical Power Plant		**Subcritical Power Plant**	
	Case 9	**Case 10**	**Case 9**	**Case 10**
GROSS (STEAM TURBINE) POWER at Generator Terminal, kWe[1]	**582,600**	**672,700**	**582,600**	**672,698**
AUXILIARY LOAD SUMMARY, kWe				
- Coal Handling and Conveying	450	540	450	540
- Pulverizers	2,970	4,180	2,970	4,180
- Sorbent Handling & Reagent Preparation	950	1,370	950	1,370
- Ash Handling	570	800	570	800
- Primary Air Fans	1,400	1,960	1,400	1,961
- Forced Draft Fans	1,780	2,500	1,780	2,500
- Induced Draft Fans	7,540	12,080	7,545	12,082
- SCR	50	70	50	70
- Baghouse	70	100	70	100
- Wet FGD	3,180	4,470	3,180	4,470
- Econamine FG Plus Auxiliaries	-	22,400	-	22,400
- CO2 Compression	-	48,790	-	48,790
- Miscelleaneous Balance of Plant	2,000	2,000	2,000	2,000
- Steam Turbine Auxiliaries	400	400	400	400
- Condensate Pumps	890	700	890	700
- Circulating Water Pumps[2]	5,250	11,190	5,247	4,159 (Power) + 7,031 (Capture)
- Ground Water Pumps	530	1,020	530	1,020
- Cooling Tower Fans[2]	2,720	5,820	2,719	2,155 (Power) + 3,665 (Capture)
- Transformer Losses	1,830	2,350	1,830	2,350
TOTAL AUXILIARIES, kWe	**32,580**	**122,740**	**32,582**	**122,742**
NET POWER, kWe	**550,020**	**549,960**	**550,018**	**549,956**
- Net Plant Efficiency (HHV)	36.8%	26.2%	36.8%	26.2%
- Net Plant Heat Rate, BTU/kWh	9,277	13,046	9,277	13,046
CONDENSER COOLING DUTY (MMBTU/hr)[3]	**2,432**	**1,928**	**2,431**	**1,927**

	Bituminous Baseline Study (DOE/NETL, 2010)		Process Validation	
	Subcritical Power Plant		Subcritical Power Plant	
	Case 9	Case 10	Case 9	Case 10
CONSUMABLES				
- As-Received Coal Feed, lb/hr	437,378	614,994	437,378	614,994
- HHV Thermal Input, kW	1,495,379	2,102,643	1,495,381	2,102,645
FLUE GAS				
- Flow Rate before Baghouse, lb/hr	4,814,113	6,755,062	4,814,607	6,755,062
- Temperature before Baghouse, lb/hr	337	337	336.9	336.8
- Composition (%by mole)				
- Ar	0.87%	0.87%	0.87%	0.87%
- CO_2	14.47%	14.50%	14.51%	14.54%
- H_2	0.00%	0.00%	0.00%	0.00%
- H_2O	8.68%	8.70%	8.72%	8.73%
- N_2	73.25%	73.24%	73.24%	73.23%
- O_2	2.50%	2.47%	2.46%	2.42%
- SO_2	0.21%	0.21%	0.21%	0.21%
STEAM EXTRACTION FROM IP/LP TURBINE TO CAPTURE PLANT				
- Flow Rate, lb/hr	-	1,931,497	-	1,931,497
- Temperature, °F	-	564.7	-	563
- Pressure, psia	-	73.5	-	73.5

Notes:

1. Boiler feed pump is connected to boiler feed pump turbine drive. Load of this pump is not included in the gross power at generator terminal.
2. Condenser heat duty for Case 9 (2432 MMBtu/hr) is higher than Case 10 (1,928 MMBtu/hr); however, the cooling tower fan load for Case 9 (2,720 kW) is lower than Case 10 (5,820 kW). It is assumed that Case 10 has already included the fan load for both Power Island and Capture Island.
3. Condenser cooling duty shown in the Table is based on a steam cycle only.

Appendix C PERFORMANCE AND ECONOMIC CALCULATIONS

Calculation Sheet

Customer ADA-ES
Project Title Solid Sorbent CO_2 Capture Pilot System
Calculation Title Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant
Elec File Location
Project File Location V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx

INITIAL DESIGN / DOE INDIRECT COSTS

Proj No 111100208
Calc No
Phase/CTR

Rev	**Date**	**By**	**Checked**	**Rev**	**Date**	**By**	**Checked**	**Rev**	**Date**	**By**	**Checked**	**Rev**	**Date**	**By**	**Checked**	**Rev**	**Date**	**By**	**Checked**
Elec File Location	26-Nov-15	BT	MER																

Line No.	**Description**	**unit**	**CASE 9 - (REF. PP)**	**CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)**	**CASE 10 WITH POLYSTYRENE BN SORBENT**	**NOTE**
1	**PLANT PERFORMANCE**					
2	A. Gross Power Output	kWe	582,600	672,700	718,738	
3						
4	B. Parasitic Power					
5	· Auxiliary Power Requirement					
6	Power Plant + FGD					
7	Coal Handling and Conveying	kWe	450	540	612.00	
8	Pulverizers	kWe	2,970	4,180	4,738	
9	Sorbent Handling & Reagent Prepara	kWe	950	1,370	1,553	
10	Ash Handling	kWe	570	800	907	
11	Primary Air Fans	kWe	1,400	1,960	2,222	
12	Forced Draft Fans	kWe	1,780	2,500	2,833	
13	Induced Draft Fans	kWe	7,540	12,080	13,671	
14	SCR	kWe	50	70	79	
15	Baghouse	kWe	70	100	113	
16	Wet FGD	kWe	3,180	4,470	5,066	
17	Miscellaneous Balance of Plant	kWe	2,000	2,000	2,000	
18	Steam Turbine Auxiliaries	kWe	400	400	400	
19	Condensate Pumps	kWe	890	700	621	
20	Circulating Water Pump	kWe	5,250	11,190	3,703	
21	Ground Water Pumps	kWe	530	1,020	775	
22	Cooling Tower Fans	kWe	2,720	5,820	1,918	
23	Transformer Losses	kWe	1,830	2,350	2,511	
24	**Sub-Total**	**kWe**	**32,580**	**51,550**	**43,722**	
25						
26	Flue Gas Polishing Scrubber Unit					
27	Recirculation Pump	kWe	0	0	1,138	
28	Recirculation Pump	kWe	0	0	1,138	
29	Bleed Pump	kWe	0	0	284	
30	Bleed Pump	kWe	0	0	284	
31	**Sub-Total**	**kWe**	**0**	**0**	**2,845**	
32						
33	CO_2 Capture					
34	Econamine FG Plus Auxiliaries (including Flue Gas Supply, SO_2 polishing, CO_2 abs. Solvent stripping and reclaiming)	kWe	0	22,400	0	
35	Condensate Return Pump	kWe	0	0	325	68)
36	Cooling Tower Fan Power	kWe	0	0	2,503	83) / CO_2 removal rate
37	Circulating Water Pump	kWe	0	0	4,802	83) / CO_2 removal rate
38	Flue Gas Compressor	kWe	0	0	53,617	Scaling CO_2 Removal Rate line (Line: 63)
39	CO_2 Recycle Compressor	kWe	0	0	14,787	Scaling CO_2 Removal Rate line (Line: 63)
40	Lean Sorbent Lift Compressor (NEW)	kWe	0	0	16,527	Scaling CO_2 Removal Rate line (Line: 63)
41	**Sub-Total**	**kWe**	**0**	**22,400**	**92,562**	
42						
43	CO_2 Compressor					
44	CO_2 Compressor	kWe	0	48,790	62,165	
45	**Sub-Total**	**kWe**	**0**	**48,790**	**62,165**	
46	**Total Auxiliary Power**	**kWe**	**32,580**	**122,740**	**201,293**	
47						
48	**GRAND Total Parasitic Power**	**kWe**	**32,580**	**122,740**	**201,293**	
49						
50	C. Back Pressure Turbine Generator Outp	kWe	0	0	32,525	
51	D. Net Power Output	kWe	550,020	549,960	549,970	
52	E. Net Power Reduction	kWe	N/A	0	0	
53	F. Coal HHV	Btu/lbm	11,666	11,666	11,666	
54	G. Coal Flow Rate	lbm/hr	437,378	614,994	697,032	
55	H. Thermal Input	kW_{th}	1,495,379	2,102,644	2,383,129	
56	· Power Plant	kW_{th}	1,495,379	1,513,467		
57	· CO_2 Capture	kW_{th}	N/A	589,177		
58	I. Gross Output to Thermal Input to PP	%	39.0	44.4	44.4	

Calculation Sheet

Customer ADA-ES
Project Title Solid Sorbent CO_2 Capture Pilot System
Calculation Title Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant
Elec File Location
Project File Location V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx

INITIAL DESIGN / DOE INDIRECT COSTS

Proj No 111100208
Calc No
Phase/CTR

Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked
Elec File Location	26-Nov-15	BT	MER																

Line No.	Description	unit	CASE 9 - (REF. PP)					CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)					CASE 10 WITH POLYSTYRENE BN SORBENT					NOTE
55	J. Net Plant HHV Efficiency	%					36.8					26.2					23.1	
56	K. Energy Penalty	%					N/A					10.6					13.7	
57	L. Net Plant HHV Heat Rate	Btu/kWh					9,277					13,046					14,785	
58	M. Capacity Factor	%					85					85					85	
59	N. Net kWh @ 85% Capacity Factor	kWh-net					4,095,448,920					4,095,002,160					4,095,075,837.00	
60																		
61	**CO_2 CAPTURE PROFILE**																	
62	A. CO_2 Produced	lbm/hr					1,038,691					1,459,838					1,651,831.00	
63	B. CO_2 Captured	lbm/hr					0					1,313,960					1,486,648	
64	C. CO_2 Emission	lbm/hr					1,038,691					145,878					165,183	
65																		
66	**CO_2 REMOVAL SYSTEM PARAMETERS**																	
67	A. Solvent Regeneration Energy	Btu/lbm CO_2					N/A					1,530					2,042	
68	B. Total Energy Requirement	MMBtu/hr					N/A					2,010					3,036	
69	· *Heat of Desorption*	*MMBtu/hr*					*N/A*					*2,010*						
70	· *Sensible Heat of Solid / Liquid*	*MMBtu/hr*					*N/A*					*0*						
71	- Heat Capacity of Solid	(Btu/lbm °F)					N/A					0						
72	- Delta Temp	(°F)					N/A					0						
73	- Solid/liquid recirc. Rate	(lbm/hr)					N/A					0						
74	· *Sensible Heat of CO_2*	*MMBtu/hr*					*N/A*					*0*						
75	- Heat Capacity of CO_2	(Btu/lbm°F)					N/A					0						
76	- Delta Temp	(°F)					N/A					0						
77	C. Steam Pressure	psia					N/A					74					45	
78	D. Steam Temperature	°F					N/A					306					274	
79	J. Steam Extraction Flow	lbm/hr					N/A					1,931,497					2,709,721	
80																		
81	**CIRC. COOLING WATER RATE**																	
82	A. CW Requirement for Power Plant	gpm					170,000					200,000					226,679	
83	B. CW Requirement for Carbon Capture	gpm					0					340,000					260,441	Scaling CO_2 Removal Rate line((Line: 113)
84	**Total**	**gpm**					**170,000**					**540,000**					**487,120**	
85																		
86	**CW PARAMETERS**																	
87	A. Cooling Water to condenser	°F					60					60					60	DOE's Report - 2010/1397, p.327
88	B. Cooling Water from condenser	°F					80					80					80	DOE's Report - 2010/1397, p.327
89	C. Cooling Water Temp. range	°F					20					20					20	DOE's Report - 2010/1397, p.42
90	D. Cooling Water Makeup Rate	gpm																
91	· Evaporative losses	%					0.8					0.8					0.8	% of Circ. Water flow rate/10°F of range, DOE's Report-2010/1397, p.42
92	· Drift losses	%					0.001					0.001					0.001	% of Circ. Water flow rateDOE's Report - 2010/1397, p.42
93	· Blowdown losses	-					= Evaporative losses / (Cycles of Concentration - 1)					= Evaporative losses / (Cycles of Concentration - 1)					= Evaporative losses / (Cycles of Concentration - 1)	DOE's Report - 2010/1397, p.42
94	; Cycles of concentration	-					4					4					4	DOE's Report - 2010/1397, p.42
95																		
96	**WATER DEMAND FOR COOLING TOWER**																	
97	A. Calculated Water Demand																	Assume the CW makeup flow rate (p.42) is equal to the water demand.
98	· Evaporative losses	gpm					2,720					8,640					7,794	
99	· Drift losses	gpm					1.7					5.4					4.9	
100	· Blowdown losses	gpm					906.7					2,880					2,598	
101	**Total**	**gpm**					**3,628.4**					**11,525**					**10,397**	
102	B. DOE's Water Demand	gpm					**5,404**					**11,580**					**N/A**	DOE's Report - 2010/1397, p.333 for Case 9 and p.359 for Case 10
103	C. % Error	%					33					0.5					N/A	
104																		
105	**RAW WATER CONSUMPTION**		(Exhibit 4-11, p.333)					(Exhibit 4-24, p.359)					(Calculation)					
106	A. RW Consumption		Demand	Internal	Withdraw.	Discharge	Consumption	Demand	Internal	Withdraw.	Discharge	Consumption	Demand	Internal	Withdraw.	Discharge	Consumption	
107	· Capture Makeup	gpm	0	0	0	0	0	39	0	39	0	39	0	0	0	0	0	Assumed
108	· FGD Makeup	gpm	1,017	0	1,017	0	1,017	1,460	0	1,460	0	1,460	2,052	0	2,052	0	2,052	Scaling coal flow rate * 1.24 for Demand; Assume zero for Int. & Withdrav
109	· BFW Makeup	gpm	74	0	74	0	74	104	0	104	0	104	146	0	146	0	14,032	Scaling coal flow rate * 1.24 for Demand; Assume zero for Int. & Withdrav
110	· Cooling Tower	gpm	5,404	600	4,804	1,215	3,589	11,580	1,959	9,621	2,604	7,017	10,397	1,758.83	8,638	2,337.93	6,300	Scaling Case 10 - Internal and Discharge.
111	**Total**	**gpm**	**6,495**	**600**	**5,895**	**1,215**	4,680	**13,183**	**1,959**	**11,224**	**2,604**	8,620	**12,595**	**1,759**	**10,836**	**2,338**	22,384	
112		**(1000 gal/D)**	**9,353**	**864**	**8,489**	**1,750**	6,739	**18,983.52**	**2,820.96**	**16,163**	**3,750**	12,413	**18,137**	**2,533**	**15,604**	**3,367**	12,237	
113	B. 50% RW Withdrawal (Calculated	(1000 gal/D)					**4,244**					**8,081**					**7,802**	
114	C. DOE's RW Consumption	(1000 gal/D)	(Exhibit 4-17, p.352)				**4,245**	(Exhibit 4-30, p.379)				**8,081**					N/A	Daily consumption is calculated on a 100% operating capacity basis.
115	D. % Error	%					0.0					0.0					N/A	

Calculation Sheet

Customer	ADA-ES	**Proj No**	111100208
Project Title	Solid Sorbent CO_2 Capture Pilot System	**Calc No**	
Calculation Title	Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant	**Phase/CTR**	
Elec File Location			
Project File Location	V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx		

INITIAL DESIGN / DOE INDIRECT COSTS

Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked
Elec File Location	26-Nov-15	BT	MER																

Line No.	Description	unit	CASE 9 - (REF. PP)				CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)				CASE 10 WITH POLYSTYRENE BN SORBENT		NOTE
116													
117	**COSTS**												
118	**A. Capital Cost**		June	2007	Oct	2015	June	2007	Oct	2015	Oct	2015	
119	· Bare Erected Cost (BEC)			$/kW	**14.15**	$/kW		$/kW	**14.15**	$/kW	$ 1,136,090,912	$/kW	Annual inflation rate is 14.15%.
120	Acct 1. Coal & Sorbent Handling		$ 31,895,000	58	$ 36,408,784	66	$ 39,542,000	72	$ 45,137,989	82	$ 48,965,507	89	Scaling coal flow rate linearly (Line: 54)
121	Acct 2. Coal & Sorbent Prep & Feed		$ 15,075,000	27	$ 17,208,416	31	$ 18,984,000	35	$ 21,670,618	39	$ 23,508,199	43	Scaling coal flow rate linearly (Line: 54)
122	Acct 3. Feedwater & Misc. BOP Syster		$ 58,766,000	107	$ 67,082,571	122	$ 78,063,000	142	$ 89,110,485	162	$ 96,666,694	176	Scaling coal flow rate linearly (Line: 54)
123	Acct 4. PC Boiler		$ 221,527,000	403	$ 252,877,528	460	$ 280,980,000	511	$ 320,744,323	583	$ 347,942,144	633	Scaling coal flow rate linearly (Line: 54)
124	Acct 5. Flue Gas Cleanup		$ 112,287,000	204	$ 128,177,870	233	$ 144,350,000	262	$ 164,778,429	300	$ 178,750,973	325	Scaling coal flow rate linearly (Line: 54)
125	Acct 6. Combustion Turbine/Accesso		$ -	0	$ -	-	$ -	0	$ -	-	$ -	0	Scaling coal flow rate linearly (Line: 54)
126	Acct 7. HRSG, Ducting & Stack		$ 31,679,000	58	$ 36,162,216	66	$ 33,792,000	61	$ 38,574,248	70	$ 41,845,188	76	Scaling coal flow rate linearly (Line: 54)
127	Acct 8. Steam Turbine Generator		$ 93,508,000	170	$ 106,741,263	194	$ 105,470,000	192	$ 120,396,127	219	$ 130,605,232	237	Scaling coal flow rate linearly (Line: 54)
128	Acct 9. Cooling Water System		$ 32,189,000	59	$ 36,744,391	67	$ 52,626,000	96	$ 60,073,638	109	$ 65,167,639	118	Scaling coal flow rate linearly (Line: 54)
129	Acct 10. Ash/Spent Sorbent Handling		$ 10,833,000	20	$ 12,366,087	22	$ 13,088,000	24	$ 14,940,215	27	$ 16,207,085	29	Scaling coal flow rate linearly (Line: 54)
130	Acct 11. Accessory Electric Plant		$ 42,692,000	78	$ 48,733,777	89	$ 68,316,000	124	$ 77,984,089	142	$ 84,596,824	154	Scaling coal flow rate linearly (Line: 54)
131	Acct 12. Instrumentation & Control		$ 17,451,000	32	$ 19,920,668	36	$ 20,024,000	36	$ 22,857,799	42	$ 24,796,048	45	Scaling coal flow rate linearly (Line: 54)
132	Acct 13. Improvements to Site		$ 10,679,000	19	$ 12,190,293	22	$ 12,006,000	22	$ 13,705,091	25	$ 14,867,227	27	Scaling coal flow rate linearly (Line: 54)
133	Acct 14. Buildings & Structures		$ 45,727,000	83	$ 52,198,291	95	$ 50,207,000	91	$ 57,312,301	104	$ 62,172,152	113	Scaling coal flow rate linearly (Line: 54)
134	CO_2 Removal System		$ -	0	$ -	-	$ 284,795,000	518	$ 325,099,223	591	$ 2,139,215,668	3,890	Back pressure turbine included and Stantec-Shaw indirect costs included
135	CO_2 Compression & Drying		$ -	0	$ -	-	$ 38,060,000	69	$ 43,446,256	79	$ 102,688,141	187	
136	**Total BEC**		**$ 724,308,000**	1,317	**$ 826,812,155**	1,503	**$ 1,240,303,000**	2,255	**$ 1,415,830,829**	2,574	**$ 3,377,994,721**	6,142	
137													
138	· ECM + Home Office Expenses + Fee		$ 68,082,000	124	$ 77,716,973	141	$ 117,071,000	213	$ 133,638,902	243	$ 318,845,653.02	580	9.4% of TPC
139	· Process Contingencies (20% of CO_2 Removal System)		$ -	-	$ -	-	$ 57,960,000	105	$ 66,162,506	120	$ 435,362,067.83	792	20% of the CO_2 Removal System
140	· Project Contingencies (15 - 30% of BEC + EPC + Process Contingency)		$ 100,043,000	182	$ 114,201,097	208	$ 203,025,000	369	$ 231,757,122	421	$ 592,750,828	1,078	Project Contingencies is about 14.34% of the BEC + ECM + Process Contigency.
141	**Total Plant Cost**		**$ 892,433,000**	1,623	**$ 1,018,730,225**	1,852	**$ 1,618,359,000**	2,943	**$ 1,847,389,360**	3,359	**$ 4,724,953,270**	8,591	
142													
143	· Preproduction Costs		$ 28,543,000	52	$ 32,582,409	59	$ 48,733,000	89	$ 55,629,700	101	$ 151,119,850	275	Scaling (Preproductio/TPC)$_{\text{case 9}}$ ~3%
144	· Inventory Capital		$ 18,287,000	33	$ 20,874,978	38	$ 28,281,000	51	$ 32,283,331	59	$ 96,819,840	176	Scaling (Inv. Ca./TPC)$_{\text{case 9}}$ ~2%
145	· Initial Cost for Catalyst and Chemicals		$ -	-	$ -	-	$ 2,712,000	5	$ 3,095,803	6	$ 180,802,311	329	
146	· Land		$ 900,000	2	$ 1,027,368	2	$ 900,000	2	$ 1,027,368	2	$ 1,027,368	2	Equal to Case 10
147	· Other Owner's Costs		$ 133,865,000	243	$ 152,809,591	278	$ 242,754,000	441	$ 277,108,575	504	$ 708,743,255	1,289	Scaling (Other Own./TPC)$_{\text{case 9,10}}$ ~15%
148	· Financing Costs		$ 24,096,000	44	$ 27,506,069	50	$ 43,696,000	79	$ 49,879,863	91	$ 127,575,374	232	Scaling (Fin. Cost/TPC)$_{\text{case 9 ,10}}$~27%
149	**Total Overnight Cost**		**$ 1,098,124,000**	1,997	**$ 1,253,530,640**	2,279	**$ 1,985,435,000**	3,610	**$ 2,266,413,999**	4,121	**$ 5,991,041,269**	10,893	
150													
151	TASC Multiplier (IOU, 35 Years)		1.134	Low Risk			1.140	High Risk			1.140	High Risk	
152	**Total As-Spent Cost (TASC)**		**$ 1,245,272,616**	2,264	**$ 1,421,503,746**	2,584	**$ 2,263,395,900**	4,115	**$ 2,583,711,959**	4,698	**$ 6,829,787,046**	12,417	Equal to TOC + Interest
153													
154	**B. Operating & Maintence Labor**												
155	B.1 Total Fixed Operating Costs												
156	· Annual Operating Labor Cost		$ 5,524,319	10	$ 6,306,121	11	$ 6,444,907	12	$ 7,356,991	13	$ 7,356,991	13	Equal to Case 10
157	· Maintenance Labor Cost		$ 5,842,145	11	$ 6,668,926	12	$ 10,429,543	19	$ 11,905,533	22	$ 31,471,100	57	Scaling (Main. Labor Cost/TPC)$_{\text{case 9,10}}$ ~0.5%
158	· Administrative & Support Labor		$ 2,841,616	5	$ 3,243,762	6	$ 4,218,612	8	$ 4,815,630	9	$ 4,815,630	9	Equal to Case 10
159	· Property Taxes and Insurance		$ 17,848,664	32	$ 20,374,609	37	$ 32,367,148	59	$ 36,947,751	67	$ 97,667,725	178	Scaling (Prop. Tax/TPC)$_{\text{case 9,10}}$ ~1.6%
160	**Total Fixed Operating Costs**		**$ 32,056,744**	58	$ 36,593,418	67	**$ 53,460,210**	97	**$ 61,025,905**	111	**$ 141,311,446**	257	
161													

Line No.	Description	unit	CASE 9 - (REF. PP)						CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)						CASE 10 WITH POLYSTYRENE BN SORBENT			NOTE
162	B.2 Total Variable Operating Costs		**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	The annual cost is calculated on a 85% operating capacity (see p.52).
163	· Maintenance Material Cost		$ -	$ 8,763,218	0.00214	$ -	$10,003,390	0.00244	$ -	$ 15,644,314	0.00382	$ -	$17,858,299	0.00436	$ -	$ 47,206,647.81	0.01153	Scaling (Main. Mat Cost/TPC)$_{\text{case 9,10}}$ ~0.7%
164	· Consumables																	
165	- Water		$ -	$ 1,424,619	0.00035	$ -	$ 1,626,231	0.00040	$ -	$ 2,711,996	0.00066	$ -	$ 3,095,798	0.00076	$ -	$ 2,988,883	0.00073	Use DOE's price (p.379) ~ 0.00115 $/US gallon
	Daily water consumption @ 100% CF	(1000 gal/D)		4,245			4,245			8,081			8,081			7,802		
	Annual water consumption @ 85% CF	(1000 gal/year)		1,317,048			1,317,048			2,507,130			2,507,130			2,420,545.28		
	Cost of Water	($/1000 gal)		$ 1.08			$ 1.23			$ 1.08			$ 1.23			$ 1.23		
166	- Chemicals																	
167	MU & WT Chem.		$ -	$ 1,103,371	0.00027	$ -	$ 1,259,520	0.00031	$ -	$ 2,100,447	0.00051	$ -	$ 2,397,703	0.00059	$ -	$ 6,338,089.47	0.00155	Scaling (MU&WT/TPC)$_{\text{case 9,10}}$ ~0.1%
168	Limestone (FGD)		$ -	$ 3,496,290	0.00085	$ -	$ 3,991,085	0.00097	$ -	$ 5,043,346	0.00123	$ -	$ 5,757,081	0.00141	$ -	$ 6,525,052	0.00159	Scaling coal flow rate linearly (Line: 54)
169	Carbon (Mercury Removal)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	
170	MEA Solvent (MEA Carbon Capture)		$ -	$ -	-	$ -	$ -	-	$ 2,513,263	$ 1,105,563	0.00027	$ 2,868,940	$ 1,262,022	0.00031	$ -	$ -	-	
171	NaOH (SO_2 Polishing Scrubber - MEA)		$ -	$ -	-	$ -	$ -	-	$ 34,221	$ 1,061,704	0.00026	$ 39,064	$ 1,211,956	0.00030	$ 44,275	$ 2,747,253	0.00067	Scaling coal flow rate (Case 10) linearly (x2 to get to 1 ppm)
172	H_2SO_4		$ -	$ -	-	$ -	$ -	-	$ 10,450	$ 324,217	0.00008	$ 11,929	$ 370,100	0.00009	$ 13,520	$ 419,470	0.00010	Scaling coal flow rate linearly (Line: 54)

Calculation Sheet

Customer ADA-ES
Project Title Solid Sorbent CO_2 Capture Pilot System
Calculation Title Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant
Elec File Location
Project File Location V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx

INITIAL DESIGN / DOE INDIRECT COSTS

Proj No 111100208
Calc No
Phase/CTR

Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked
Elec File Location	26-Nov-15	BT	MER																				

Line No.	Description	unit	CASE 9 - (REF. PP)						CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)						CASE 10 WITH POLYSTYRENE BN SORBENT			NOTE
173	Corrosion Inhibitor (MEA Carbon Capture)		$ -	$ -	-	$ -	$ -	-	$ 154,511	$ 7,358	0.00000	$ 176,377	$ 8,399	0.00000	$ -	$ -	-	
174	Activated Carbon (MEA Carbon Capture)		$ -	$ -	-	$ -	$ -	-	$ -	$ 616,433	0.00015	$ -	$ 703,671	0.00017	$ -	$ -	-	
175	Ammonia (19% NH_3) (SCR)		$ -	$ 3,136,289	0.00077	$ -	$ 3,580,137	0.00087	$ -	$ 4,446,378	0.00109	$ -	$ 5,075,630	0.00124	$ -	$ 5,752,698	0.00140	Scaling coal flow rate linearly (Line: 54)
176	Solid Sorbent		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ 180,744,516	$ 55,610,859	0.01358	¥ % attrition rate (0.005%) and CO_2 removal rate
	Initial fill														Total mass, lb (44 trains)			
	Adsorber-1														169,450			
	Adsorber-2														134,562			
	Adsorber-3														383,401			
	Regenerator														69,700			
	Total	(lbm)													757,112			
	Total for 44 Trains	(lbm)													33,312,929			
	Grand Total	(Ton)													**16,656**			
	Cost of Sorbent	($/ton)													$ 10,851			
	Solid Comsumption (Mainly Attrition)																	
	%Attrition	(% / Cycle)														0.005		
	Total Sorbent flow rate	(lbm/hr)														27,530,510		The solid circ. rate linearly depends on the CO_2 captured flow rate.
	Spent solid	(lbm/hr)														1,377		
	Daily solid consumption @ 100% CF	(ton/day)														16.52		
	Annual solid consumption @ 85% CF	(ton/year)														5,125		
177	NaOH (CT - Solid CC)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 220,007	0.00005	Scaling CW requirement for CO_2 capture (83)
	Old CW recir. Rate	(gpm)														27,300		
	Old NaOH requirement	(lbm/day)														300.30		
	New NaOH requirement	(lbm/day)														2,864.85		
	Daily NaOH consumption @ 100% CF	(ton/day)														1.43		
	Annual NaOH consumption @ 85% CF	(ton/year)														444.41		
	Cost of NaOH	($/ton)														495.05		
178	Lime (CT - Solid CC)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 694,431	0.00017	Scaling CW requirement for CO_2 capture (83)
	Old CW recir. Rate	(gpm)														27,300		
	Old Lime requirement	(lbm/day)														1,965.60		
	New Lime requirement	(lbm/day)														18,751.76		
	Daily Lime consumption @ 100% CF	(ton/day)														9.38		
	Annual Lime consumption @ 85% CF	(ton/year)														2,908.87		
	Cost of Lime	($/ton)														238.73		
179	H_2SO_4 (CT - Solid CC)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 236,812	0.00006	Scaling CW requirement for CO_2 capture (83)
	Old CW recir. Rate	(gpm)														27,300		
	Old H_2SO_4 requirement	(lbm/day)														1,010.10		
	New H_2SO_4 requirement	(lbm/day)														9,636.32		
	Daily H_2SO_4 consumption @ 100% CF	(ton/day)														4.82		
	Annual H_2SO_4 consumption @ 85% CF	(ton/year)														1,494.83		
	Cost of H_2SO_4	($/ton)														158.42		
180	**Subtotal Chemicals**		**$ -**	**$ 7,735,950**	0.00189	**$ -**	**$ 8,830,743**	0.00216	**$ 2,712,445**	**$ 14,705,446**	0.00359	**$ 3,096,311**	**$16,786,562**	0.00410	**$ 180,802,311**	**$ 78,544,671**	0.01918	
181	- Other																	
182	Supplemental Fuel		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	
183	SCR Catalyst		$ -	$ 592,641	0.00014	$ -	$ 676,512	0.00017	$ -	$ 831,516	0.00020	$ -	$ 949,192	0.00023	$ -	$ 1,075,811	0.00026	Scaling coal flow rate linearly (Line: 54)
184	Emission Penalties		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	
185	**Subtotal Other**		**$ -**	**$ 592,641**	0.00014	**$ -**	**$ 676,512**	0.00017	**$ -**	**$ 831,516**	0.00020	**$ -**	**$ 949,192**	0.00023	**$ -**	**$ 1,075,811**	0.00026	
186	- Waste Disposal																	
187	Fly Ash		$ -	$ 2,049,540	0.00050	$ -	$ 2,339,591	0.00057	$ -	$ 2,881,846	0.00070	$ -	$ 3,289,685	0.00080	$ -	$ 3,728,516	0.00091	Scaling coal flow rate linearly (Line: 54)
188	Bottom Ash		$ -	$ 512,385	0.00013	$ -	$ 584,898	0.00014	$ -	$ 720,462	0.00018	$ -	$ 822,422	0.00020	$ -	$ 932,130	0.00023	Scaling coal flow rate linearly (Line: 54)
189	Solid Waste		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 94,947	0.00002	Assume 0.005% Attrition per cycle for 27,530,510 lbm solid/hr
	Solid Waste equal to Solid Consumption																	
	Cost of Solid Waste Treament	($/ton)														$ 18.53		
190	Waste water		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	Assume wastewater cost is accounted in the chemical costs.
191	**Subtotal Waste Disposal**		**$ -**	**$ 2,561,925**	0.00063	**$ -**	**$ 2,924,489**	0.00071	**$ -**	**$ 3,602,308**	0.00088	**$ -**	**$ 4,112,107**	0.00100	**$ -**	**$ 4,755,592**	0.00116	
192	- By-products & Emissions																	
193	Gypsum		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	Cost of Gypsum Byproduct is zero.
194	**Subtotal By-Product**		**$ -**	**$ -**	-	**$ -**	**$ -**	-	**$ -**	**$ -**	-	**$ -**	**$ -**	-	**$ -**	**$ -**	-	
195	**Grand Total Variable Operating Costs**		**$ -**	**$21,078,353**	0.00515	**$ -**	**$24,061,364**	0.00588	**$ 2,712,445**	**$ 37,495,580**	0.00916	**$ 3,096,311**	**$42,801,959**	0.01045	**$ 180,802,311**	**$ 134,571,604**	0.03286	
196																		
197	B.3 Fuel (Variable Operating Costs)																	
198	· Coal (for Power Plant)		$ -	$62,175,575	0.01518	$ -	$70,974,670	0.01733	$ -	$ 87,425,787	0.02135	$ -	$99,798,295	0.02437	$ -	$ 113,110,975	0.02762	Calculated from coal feed rate with DOE's cost
	Coal flow rate	(lbm/hr)		437,378			437,378			614,994			614,994			697,032		

Stantec

Calculation Sheet

Customer ADA-ES
Project Title Solid Sorbent CO_2 Capture Pilot System
Calculation Title Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant
Elec File Location
Project File Location V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx

INITIAL DESIGN / DOE INDIRECT COSTS

Proj No 111100208
Calc No
Phase/CTR

Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked
Elec File Location	26-Nov-15	BT	MER																				

Line No.	Description	unit	CASE 9 - (REF. PP)						CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)						CASE 10 WITH POLYSTYRENE BN SORBENT			NOTE
	Daily coal consumption @ 100% CF	(ton/day)		5,249			5,249			7,380			7,380			8,364		
	Annual coal consumption @ 85% CF	(ton/year)		1,628,358			1,628,358			2,289,624			2,289,624			2,595,050		
	Cost of coal	($/ton)		$ 38.18			$ 43.59			$ 38.18			$ 43.59			$ 43.59		
199	**Total Fuel Costs**		$ -	**$62,175,575**	0.01518	$ -	**$70,974,670**	0.01733	$ -	**$ 87,425,787**	0.02135	$ -	**$99,798,295**	0.02437	$ -	**$ 113,110,975**	0.02762	
200																		
201	**C. CO_2 Transport, Storage & Monitoring**																	
202	C.1 Transport Costs		June		2007	Oct		2015	June		2007	Oct		2015	Oct		2015	
203	· Pipeline Costs		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
204	· Related Capital Expenditures		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
205	· O&M Costs		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
206	C.2 Storage Costs																	
207	· Site Screening and Evaluation		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
208	· Injection Wells		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
209	· Injection Equipment		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
210	· O&M Costs		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
211	· Pore Volume Acquisition		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
212	C.3 Monitoring Costs		$ -		-	$ -		-	$ -		-	$ -		-	$ -		-	
213	**Total CO_2 TS&M Costs**		$ -		-	$ -		-	$ 23,751,018		43	$ 27,112,265		49	$ 30,675,509		56	Scaling CO_2 Removal Rate line((Line: 63)
214																		
215	**Economic Parameters**		June		2007	Oct		2015	June		2007	Oct		2015	Oct		2015	
216	Type of Risk		Low			Low			High			High			High			
217			COE	LCOE		COE	LCOE		COE	LCOE		COE	LCOE		COE	LCOE		
218	1. Capital Charge Factor, 5 Years		0.116	0.116		0.116	0.116		0.124	0.124		0.124	0.124		0.124	0.124		
219	2. Nominal (Apparent) Escalation Rate		3%	0%		3%	0%		3%	0%		3%	0%		3%	0%		
220	3. General Inflation Rate		3%	0%		3%	0%		0%	0%		0%	0%		0%	0%		
221	4. Real Escalation Rate		0%	0%		0%	0%		0%	0%		0%	0%		0%	0%		
222	5. Debt (% of Total)		50	50		50	50		45	45		45	45		45	45		
223	6. Equity (% of Total)		50	50		50	50		55	55		55	55		55	55		
224	7. Real Discount Rate (Nominal)		8.25%	8.25%		8.25%	8.25%		9.08%	9.08%		9.08%	9.08%		9.08%	9.08%		
225	8. Real Discount Rate (After Tax)		7.39%	7.39%		7.39%	7.39%		8.13%	8.13%		8.13%	8.13%		8.13%	8.13%		
226																		
227	**COE contribution (¢/kWh)**		June		2007	Oct		2015	June		2007	Oct		2015	Oct		2015	**PROCEDURE 2 - LCOE Calculation**
228	1. Capital Costs		3.11			3.55			6.01			6.86			18.14			LCOE is obtained when the difference between the NPV of COE and
229	2. Fixed Operating Costs		0.78			0.89			1.31			1.49			3.45			LCOE is equal to zero.
230	3. Variable Operating Costs		0.51			0.59			0.92			1.05			3.29			Step 1: Change cost parameters (Lines 123 - 230).
231	4. Fuel Costs		1.52			1.73			2.13			2.44			2.76			Step 2: Use Solver in Data Tab. In the Solver's windor, set parameters
232	5. CO_2 T&SM Costs		0.00			0.00			0.58			0.66			0.75			as follows:
233	**Total COE (2007)**		**5.93**	**7.53**		**6.76**	**8.59**		**10.95**	**13.9**		**12.50**	**15.87**		**28.39**	**35.99**		- Set Target Cell: R400
234			**5.94**			**6.78**			**10.97**			**12.52**						- Select option "Value of" and set the blank cell equal to 0.
235	**Total COE (2012, @ 3% average annual inflation rate)**		6.89	8.73		7.86	9.96		12.72	16.1		14.52	18.40		32.91	41.72		- By Changing Cells: T358
236																		- Click Solve button to get new LCOE.
237	**Year**		**COE**	**LCOE**		**COE**	**LCOE**		**COE**	**LCOE**		**COE**	**LCOE**		**COE**	**LCOE**		
238	**2007** **0** ase Year - Con		(2.681)	(2.681)		(3.061)	(3.061)		(4.848)	(4.848)		(5.534)	(5.534)		(14.63)	(14.63)		
239	**2008** **1** Construction		(8.334)	(8.334)		(9.513)	(9.513)		(15.067)	(15.067)		(17.200)	(17.200)		(45.47)	(45.47)		
240	**2009** **2** Construction		(7.195)	(7.195)		(8.213)	(8.213)		(13.008)	(13.008)		(14.849)	(14.849)		(39.25)	(39.25)		
241	**2010** **3** Construction		(5.963)	(5.963)		(6.807)	(6.807)		(10.781)	(10.781)		(12.307)	(12.307)		(32.53)	(32.53)		
242	**2011** **4** Construction		(4.633)	(4.633)		(5.289)	(5.289)		(8.377)	(8.377)		(9.562)	(9.562)		(25.28)	(25.28)		
243	**2012** **5** Operating		6.886	8.728		7.861	9.963		12.72	16.120		14.52	18.401		32.91	41.720		
244	**2013** **6** Operating		7.093	8.728		8.096	9.963		13.10	16.120		14.95	18.401		33.90	41.720		
245	**2014** **7** Operating		7.305	8.728		8.339	9.963		13.49	16.120		15.40	18.401		34.92	41.720		
246	**2015** **8** Operating		7.525	8.728		8.589	9.963		13.90	16.120		15.86	18.401		35.96	41.720		
247	**2016** **9** Operating		7.750	8.728		8.847	9.963		14.31	16.120		16.34	18.401		37.04	41.720		
248	**2017** **10** Operating		7.983	8.728		9.113	9.963		14.74	16.120		16.83	18.401		38.15	41.720		
249	**2018** **11** Operating		8.222	8.728		9.386	9.963		15.18	16.120		17.33	18.401		39.30	41.720		
250	**2019** **12** Operating		8.469	8.728		9.668	9.963		15.64	16.120		17.85	18.401		40.48	41.720		
251	**2020** **13** Operating		8.723	8.728		9.958	9.963		16.11	16.120		18.39	18.401		41.69	41.720		
252	**2021** **14** Operating		8.985	8.728		10.256	9.963		16.59	16.120		18.94	18.401		42.94	41.720		
253	**2022** **15** Operating		9.254	8.728		10.564	9.963		17.09	16.120		19.51	18.401		44.23	41.720		**NOTE**
254	**2023** **16** Operating		9.532	8.728		10.881	9.963		17.60	16.120		20.09	18.401		45.56	41.720		**1. If the performance parameters and economic parameters are**
255	**2024** **17** Operating		9.818	8.728		11.207	9.963		18.13	16.120		20.70	18.401		46.92	41.720		**changed, follow Procedures 1 and 2.**
256	**2025** **18** Operating		10.112	8.728		11.544	9.963		18.67	16.120		21.32	18.401		48.33	41.720		**2. If only the performance parameters are changed, follow Procedures**
257	**2026** **19** Operating		10.416	8.728		11.890	9.963		19.23	16.120		21.96	18.401		49.78	41.720		**1 and 2.**

Calculation Sheet

Customer ADA-ES
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Project File Location V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx

INITIAL DESIGN / DOE INDIRECT COSTS

Proj No 111100208
Calc No
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Elec File Location	26-Nov-15	BT	MER																				

Line No.	Description			unit	CASE 9 - (REF. PP)						CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)						CASE 10 WITH POLYSTYRENE BN SORBENT			NOTE
258	**2027**	**20**	Operating		10.728	8.728		12.247	9.963		19.81	16.120		22.61	18.401		51.27	41.720		**3. If only the economic parameters are changed, follow Procedure 2.**
259	**2028**	**21**	Operating		11.050	8.728		12.614	9.963		20.41	16.120		23.29	18.401		52.81	41.720		**4. A change in Thermal Input affects the CO_2 Captured Flow Rate and**
260	**2029**	**22**	Operating		11.382	8.728		12.992	9.963		21.02	16.120		23.99	18.401		54.40	41.720		**Gross Output.**
261	**2030**	**23**	Operating		11.723	8.728		13.382	9.963		21.65	16.120		24.71	18.401		56.03	41.720		
262	**2031**	**24**	Operating		12.075	8.728		13.784	9.963		22.30	16.120		25.45	18.401		57.71	41.720		
263	**2032**	**25**	Operating		12.437	8.728		14.197	9.963		22.97	16.120		26.22	18.401		59.44	41.720		
264	**2033**	**26**	Operating		12.810	8.728		14.623	9.963		23.66	16.120		27.00	18.401		61.22	41.720		
265	**2034**	**27**	Operating		13.194	8.728		15.062	9.963		24.37	16.120		27.81	18.401		63.06	41.720		
266	**2035**	**28**	Operating		13.590	8.728		15.514	9.963		25.10	16.120		28.65	18.401		64.95	41.720		
267	**2036**	**29**	Operating		13.998	8.728		15.979	9.963		25.85	16.120		29.51	18.401		66.90	41.720		
268	**2037**	**30**	Operating		14.418	8.728		16.458	9.963		26.62	16.120		30.39	18.401		68.91	41.720		
269	**2038**	**31**	Operating		14.850	8.728		16.952	9.963		27.42	16.120		31.30	18.401		70.98	41.720		
270	**2039**	**32**	Operating		15.296	8.728		17.461	9.963		28.25	16.120		32.24	18.401		73.10	41.720		
271	**2040**	**33**	Operating		15.755	8.728		17.985	9.963		29.09	16.120		33.21	18.401		75.30	41.720		
272	**2041**	**34**	Operating		16.228	8.728		18.524	9.963		29.97	16.120		34.21	18.401		77.56	41.720		
273	**2042**	**35**	Operating		16.714	8.728		19.080	9.963		30.87	16.120		35.23	18.401		79.88	41.720		
274				**NPV**	**18.43**	**18.43**		**21.04**	**21.04**		**34.89**	**34.89**		**39.83**	**39.83**		**74.43**	**74.43**		
275				**D NPV**	**(0.00)**			**0.00**			**(0.00)**			**(0.00)**			**-**			
276				**IRROE**		**0.1225**			**0.1225**			**0.1224**			**0.1224**			**0.1224**		

Calculation Sheet

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Line No.	Description	unit	CASE 9 - (REF. PP)	CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)	CASE 10 WITH POLYSTYRENE BN SORBENT	NOTE
1	**PLANT PERFORMANCE**					
2	A. Gross Power Output	kWe	582,600	672,700	718,738	
3						
4	B. Parasitic Power					
5	· Auxiliary Power Requirement					
6	Power Plant + FGD					
7	Coal Handling and Conveying	kWe	450	540	612.00	
8	Pulverizers	kWe	2,970	4,180	4,738	
9	Sorbent Handling & Reagent Prepara	kWe	950	1,370	1,553	
10	Ash Handling	kWe	570	800	907	
11	Primary Air Fans	kWe	1,400	1,960	2,222	
12	Forced Draft Fans	kWe	1,780	2,500	2,833	
13	Induced Draft Fans	kWe	7,540	12,080	13,671	
14	SCR	kWe	50	70	79	
15	Baghouse	kWe	70	100	113	
16	Wet FGD	kWe	3,180	4,470	5,066	
17	Miscellaneous Balance of Plant	kWe	2,000	2,000	2,000	
18	Steam Turbine Auxiliaries	kWe	400	400	400	
19	Condensate Pumps	kWe	890	700	621	
20	Circulating Water Pump	kWe	5,250	11,190	3,703	
21	Ground Water Pumps	kWe	530	1,020	775	
22	Cooling Tower Fans	kWe	2,720	5,820	1,918	
23	Transformer Losses	kWe	1,830	2,350	2,511	
24	**Sub-Total**	**kWe**	**32,580**	**51,550**	**43,722**	
25						
26	Flue Gas Polishing Scrubber Unit					
27	Recirculation Pump	kWe	0	0	1,138	
28	Recirculation Pump	kWe	0	0	1,138	
29	Bleed Pump	kWe	0	0	284	
30	Bleed Pump	kWe	0	0	284	
31	**Sub-Total**	**kWe**	**0**	**0**	**2,845**	
32						
33	CO_2 Capture					
34	Econamine FG Plus Auxiliaries (including Flue Gas Supply, SO_2 polishing, CO_2 abs. Solvent stripping and reclaiming)	kWe	0	22,400	0	
35	Condensate Return Pump	kWe	0	0	325	68)
36	Cooling Tower Fan Power	kWe	0	0	2,503	83) / CO_2 removal rate
37	Circulating Water Pump	kWe	0	0	4,802	83) / CO_2 removal rate
38	Flue Gas Compressor	kWe	0	0	53,617	Scaling CO_2 Removal Rate line (Line: 63)
39	CO_2 Recycle Compressor	kWe	0	0	14,787	Scaling CO_2 Removal Rate line (Line: 63)
40	Lean Sorbent Lift Compressor (NEW)	kWe	0	0	16,527	Scaling CO_2 Removal Rate line (Line: 63)
41	**Sub-Total**	**kWe**	**0**	**22,400**	**92,562**	
42						
43	CO_2 Compressor					
44	CO_2 Compressor	kWe	0	48,790	62,165	
45	**Sub-Total**	**kWe**	**0**	**48,790**	**62,165**	
46	**Total Auxiliary Power**	**kWe**	**32,580**	**122,740**	**201,293**	
47						
48	**GRAND Total Parasitic Power**	**kWe**	**32,580**	**122,740**	**201,293**	
49						
50	C. Back Pressure Turbine Generator Outp	kWe	0	0	32,525	
51	D. Net Power Output	kWe	550,020	549,960	549,970	
52	E. Net Power Reduction	kWe	N/A	0	0	
53	F. Coal HHV	Btu/lbm	11,666	11,666	11,666	
54	G. Coal Flow Rate	lbm/hr	437,378	614,994	697,032	
55	H. Thermal Input	kW_{th}	1,495,379	2,102,644	2,383,129	
56	· Power Plant	kW_{th}	1,495,379	1,513,467		
57	· CO_2 Capture	kW_{th}	N/A	589,177		
58	I. Gross Output to Thermal Input to PP	%	39.0	44.4	44.4	

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Calculation Sheet

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Line No.	Description	unit	CASE 9 - (REF. PP)					CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)					CASE 10 WITH POLYSTYRENE BN SORBENT					NOTE
55	J. Net Plant HHV Efficiency	%					36.8					26.2					23.1	
56	K. Energy Penalty	%					N/A					10.6					13.7	
57	L. Net Plant HHV Heat Rate	Btu/kWh					9,277					13,046					14,785	
58	M. Capacity Factor	%					85					85					85	
59	N. Net kWh @ 85% Capacity Factor	kWh-net					4,095,448,920					4,095,002,160					4,095,075,837.00	
60																		
61	**CO_2 CAPTURE PROFILE**																	
62	A. CO_2 Produced	lbm/hr					1,038,691					1,459,838					1,651,831.00	
63	B. CO_2 Captured	lbm/hr					0					1,313,960					1,486,648	
64	C. CO_2 Emission	lbm/hr					1,038,691					145,878					165,183	
65																		
66	**CO_2 REMOVAL SYSTEM PARAMETERS**																	
67	A. Solvent Regeneration Energy	Btu/lbm CO_2					N/A					1,530					2,042	
68	B. Total Energy Requirement	MMBtu/hr					N/A					2,010					3,036	
69	*· Heat of Desorption*	*MMBtu/hr*					*N/A*					*2,010*						
70	*· Sensible Heat of Solid / Liquid*	*MMBtu/hr*					*N/A*					*0*						
71	- Heat Capacity of Solid	(Btu/lbm °F)					N/A					0						
72	- Delta Temp	(°F)					N/A					0						
73	- Solid/liquid recirc. Rate	(lbm/hr)					N/A					0						
74	*· Sensible Heat of CO_2*	*MMBtu/hr*					*N/A*					*0*						
75	- Heat Capacity of CO_2	(Btu/lbm°F)					N/A					0						
76	- Delta Temp	(°F)					N/A					0						
77	C. Steam Pressure	psia					N/A					74					45	
78	D. Steam Temperature	°F					N/A					306					274	
79	J. Steam Extraction Flow	lbm/hr					N/A					1,931,497					2,709,721	
80																		
81	**CIRC. COOLING WATER RATE**																	
82	A. CW Requirement for Power Plant	gpm					170,000					200,000					226,679	
83	B. CW Requirement for Carbon Capture	gpm					0					340,000					260,441	Scaling CO_2 Removal Rate line (Line: 113)
84	**Total**	**gpm**					**170,000**					**540,000**					**487,120**	
85																		
86	**CW PARAMETERS**																	
87	A. Cooling Water to condenser	°F					60					60					60	DOE's Report - 2010/1397, p.327
88	B. Cooling Water from condenser	°F					80					80					80	DOE's Report - 2010/1397, p.327
89	C. Cooling Water Temp. range	°F					20					20					20	DOE's Report - 2010/1397, p.42
90	D. Cooling Water Makeup Rate	gpm																
91	· Evaporative losses	%					0.8					0.8					0.8	% of Circ. Water flow rate/10°F of range, DOE's Report-2010/1397, p.42
92	· Drift losses	%					0.001					0.001					0.001	% of Circ. Water flow rateDOE's Report - 2010/1397, p.42
93	· Blowdown losses	-					= Evaporative losses / (Cycles of Concentration - 1)					= Evaporative losses / (Cycles of Concentration - 1)					= Evaporative losses / (Cycles of Concentration - 1)	DOE's Report - 2010/1397, p.42
94	; Cycles of concentration	-					4					4					4	DOE's Report - 2010/1397, p.42
95																		
96	**WATER DEMAND FOR COOLING TOWER**																	
97	A. Calculated Water Demand																	Assume the CW makeup flow rate (p.42) is equal to the water demand.
98	· Evaporative losses	gpm					2,720					8,640					7,794	
99	· Drift losses	gpm					1.7					5.4					4.9	
100	· Blowdown losses	gpm					906.7					2,880					2,598	
101	**Total**	**gpm**					**3,628.4**					**11,525**					**10,397**	
102	B. DOE's Water Demand	gpm					**5,404**					**11,580**					**N/A**	DOE's Report - 2010/1397, p.333 for Case 9 and p.359 for Case 10
103	C. % Error	%					33					0.5					N/A	
104																		
105	**RAW WATER CONSUMPTION**		(Exhibit 4-11, p.333)					(Exhibit 4-24, p.359)					(Calculation)					
106	A. RW Consumption		Demand	Internal	Withdraw.	Discharge	Consumption	Demand	Internal	Withdraw.	Discharge	Consumption	Demand	Internal	Withdraw.	Discharge	Consumption	
107	· Capture Makeup	gpm	0	0	0	0	0	39	0	39	0	39	0	0	0	0	0	Assumed
108	· FGD Makeup	gpm	1,017	0	1,017	0	1,017	1,460	0	1,460	0	1,460	2,052	0	2,052	0	2,052	Scaling coal flow rate * 1.24 for Demand; Assume zero for Int. & Withdraw
109	· BFW Makeup	gpm	74	0	74	0	74	104	0	104	0	104	146	0	146	0	14,032	Scaling coal flow rate * 1.24 for Demand; Assume zero for Int. & Withdraw
110	· Cooling Tower	gpm	5,404	600	4,804	1,215	3,589	11,580	1,959	9,621	2,604	7,017	10,397	1,758.83	8,638	2,337.93	6,300	Scaling Case 10 - Internal and Discharge.
111	**Total**	**gpm**	**6,495**	**600**	**5,895**	**1,215**	4,680	**13,183**	**1,959**	**11,224**	**2,604**	8,620	**12,595**	**1,759**	**10,836**	**2,338**	22,384	
112		**(1000 gal/D)**	**9,353**	**864**	**8,489**	**1,750**	6,739	**18,983.52**	**2,820.96**	**16,163**	**3,750**	12,413	**18,137**	**2,533**	**15,604**	**3,367**	12,237	
113	B. 50% RW Withdrawal (Calculated	(1000 gal/D)					**4,244**					**8,081**					**7,802**	
114	C. DOE's RW Consumption	(1000 gal/D)	(Exhibit 4-17, p.352)				**4,245**	(Exhibit 4-30, p.379)				**8,081**					N/A	Daily consumption is calculated on a 100% operating capacity basis.
115	D. % Error	%					0.0					0.0					N/A	

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Line No.	Description	unit	CASE 9 - (REF. PP)				CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)				CASE 10 WITH POLYSTYRENE BN SORBENT		NOTE
116													
117	**COSTS**												
118	**A. Capital Cost**		June	2007	Feb	2011	June	2007	Feb	2011	Feb	2011	
119	· Bare Erected Cost (BEC)			$/kW	6.22	$/kW		$/kW	6.22	$/kW	$ 1,057,136,133	$/kW	Annual inflation rate is 14.15%.
120	Acct 1. Coal & Sorbent Handling		$ 31,895,000	58	$ 33,878,487	62	$ 39,542,000	72	$ 42,001,039	76	$ 45,562,557	83	Scaling coal flow rate linearly (Line: 54)
121	Acct 2. Coal & Sorbent Prep & Feed		$ 15,075,000	27	$ 16,012,485	29	$ 18,984,000	35	$ 20,164,578	37	$ 21,874,452	40	Scaling coal flow rate linearly (Line: 54)
122	Acct 3. Feedwater & Misc. BOP Syster		$ 58,766,000	107	$ 62,420,542	113	$ 78,063,000	142	$ 82,917,584	151	$ 89,948,660	164	Scaling coal flow rate linearly (Line: 54)
123	Acct 4. PC Boiler		$ 221,527,000	403	$ 235,303,327	428	$ 280,980,000	511	$ 298,453,592	543	$ 323,761,249	589	Scaling coal flow rate linearly (Line: 54)
124	Acct 5. Flue Gas Cleanup		$ 112,287,000	204	$ 119,269,907	217	$ 144,350,000	262	$ 153,326,842	279	$ 166,328,338	302	Scaling coal flow rate linearly (Line: 54)
125	Acct 6. Combustion Turbine/Accesso		$ -	0	$ -	-	$ -	0	$ -	-	$ -	0	Scaling coal flow rate linearly (Line: 54)
126	Acct 7. HRSG, Ducting & Stack		$ 31,679,000	58	$ 33,649,055	61	$ 33,792,000	61	$ 35,893,458	65	$ 38,937,078	71	Scaling coal flow rate linearly (Line: 54)
127	Acct 8. Steam Turbine Generator		$ 93,508,000	170	$ 99,323,078	181	$ 105,470,000	192	$ 112,028,971	204	$ 121,528,575	221	Scaling coal flow rate linearly (Line: 54)
128	Acct 9. Cooling Water System		$ 32,189,000	59	$ 34,190,770	62	$ 52,626,000	96	$ 55,898,707	102	$ 60,638,691	110	Scaling coal flow rate linearly (Line: 54)
129	Acct 10. Ash/Spent Sorbent Handling		$ 10,833,000	20	$ 11,506,683	21	$ 13,088,000	24	$ 13,901,917	25	$ 15,080,743	27	Scaling coal flow rate linearly (Line: 54)
130	Acct 11. Accessory Electric Plant		$ 42,692,000	78	$ 45,346,931	82	$ 68,316,000	124	$ 72,564,437	132	$ 78,717,608	143	Scaling coal flow rate linearly (Line: 54)
131	Acct 12. Instrumentation & Control		$ 17,451,000	32	$ 18,536,243	34	$ 20,024,000	36	$ 21,269,253	39	$ 23,072,800	42	Scaling coal flow rate linearly (Line: 54)
132	Acct 13. Improvements to Site		$ 10,679,000	19	$ 11,343,106	21	$ 12,006,000	22	$ 12,752,629	23	$ 13,834,001	25	Scaling coal flow rate linearly (Line: 54)
133	Acct 14. Buildings & Structures		$ 45,727,000	83	$ 48,570,672	88	$ 50,207,000	91	$ 53,329,274	97	$ 57,851,381	105	Scaling coal flow rate linearly (Line: 54)
134	CO_2 Removal System		$ -	0	$ -	-	$ 284,795,000	518	$ 302,505,839	550	$ 1,990,546,844	3,619	Back pressure turbine included and Stantec-Shaw indirect costs included
135	CO_2 Compression & Drying		$ -	0	$ -	-	$ 38,060,000	69	$ 40,426,876	74	$ 95,551,636	174	
136	**Total BEC**		**$ 724,308,000**	1,317	**$ 769,351,286**	1,399	**$ 1,240,303,000**	2,255	**$ 1,317,434,998**	2,396	**$ 3,143,234,612**	5,715	
137													
138	· ECM + Home Office Expenses + Fee		$ 68,082,000	124	$ 72,315,885	131	$ 117,071,000	213	$ 124,351,415	226	$ 296,686,873.52	539	9.4% of TPC
139	· Process Contingencies (20% of CO_2 Removal System)		$ -	-	$ -	-	$ 57,960,000	105	$ 61,564,418	112	$ 405,105,760.51	737	20% of the CO_2 Removal System
140	· Project Contingencies (15 - 30% of BEC + EPC + Process Contingency)		$ 100,043,000	182	$ 106,264,477	193	$ 203,025,000	369	$ 215,650,724	392	$ 551,556,492	1,003	Project Contingencies is about 14.34% of the BEC + ECM + Process Contigency.
141	**Total Plant Cost**		**$ 892,433,000**	1,623	**$ 947,931,648**	1,723	**$ 1,618,359,000**	2,943	**$ 1,719,001,555**	3,126	**$ 4,396,583,739**	7,994	
142													
143	· Preproduction Costs		$ 28,543,000	52	$ 30,318,033	55	$ 48,733,000	89	$ 51,763,609	94	$ 140,617,491	256	Scaling (Preproductio/TPC)$_{case\ 9}$ ~3%
144	· Inventory Capital		$ 18,287,000	33	$ 19,424,232	35	$ 28,281,000	51	$ 30,039,740	55	$ 90,091,163	164	Scaling (Inv. Ca./TPC)$_{case\ 9}$ ~2%
145	· Initial Cost for Catalyst and Chemicals		$ -	-	$ -	-	$ 2,712,000	5	$ 2,880,654	5	$ 168,237,113	306	
146	· Land		$ 900,000	2	$ 955,969	2	$ 900,000	2	$ 955,969	2	$ 955,969	2	Equal to Case 10
147	· Other Owner's Costs		$ 133,865,000	243	$ 142,189,800	259	$ 242,754,000	441	$ 257,850,393	469	$ 659,487,807	1,199	Scaling (Other Own./TPC)$_{case\ 9,10}$ ~15%
148	· Financing Costs		$ 24,096,000	44	$ 25,594,483	47	$ 43,696,000	79	$ 46,413,368	84	$ 118,709,283	216	Scaling (Fin. Cost/TPC)$_{case\ 9\ ,10}$~27%
149	**Total Overnight Cost**		**$ 1,098,124,000**	1,997	**$ 1,166,414,166**	2,121	**$ 1,985,435,000**	3,610	**$ 2,108,905,287**	3,835	**$ 5,574,682,566**	10,136	
150													
151	TASC Multiplier (IOU, 35 Years)		1.134	Low Risk			1.140	High Risk			1.140	High Risk	
152	**Total As-Spent Cost (TASC)**		**$ 1,245,272,616**	2,264	**$ 1,322,713,664**	2,405	**$ 2,263,395,900**	4,115	**$ 2,404,152,027**	4,372	**$ 6,355,138,125**	11,554	Equal to TOC + Interest
153													
154	**B. Operating & Maintence Labor**												
155	B.1 Total Fixed Operating Costs												
156	· Annual Operating Labor Cost		$ 5,524,319	10	$ 5,867,866	11	$ 6,444,907	12	$ 6,845,703	12	$ 6,845,703	12	Equal to Case 10
157	· Maintenance Labor Cost		$ 5,842,145	11	$ 6,205,456	11	$ 10,429,543	19	$ 11,078,136	20	$ 29,283,956	53	Scaling (Main. Labor Cost/TPC)$_{case\ 9,10}$ ~0.5%
158	· Administrative & Support Labor		$ 2,841,616	5	$ 3,018,330	5	$ 4,218,612	8	$ 4,480,959	8	$ 4,480,959	8	Equal to Case 10
159	· Property Taxes and Insurance		$ 17,848,664	32	$ 18,958,637	34	$ 32,367,148	59	$ 34,379,997	63	$ 90,880,122	165	Scaling (Prop. Tax/TPC)$_{case\ 9,10}$ ~1.6%
160	**Total Fixed Operating Costs**		**$ 32,056,744**	58	$ 34,050,290	62	**$ 53,460,210**	97	**$ 56,784,795**	103	**$ 131,490,741**	239	
161													

Line No.	Description	unit	Case 9 Initial Cost	Case 9 Annual Cost	$/kWh-net	Case 9 Initial Cost	Case 9 Annual Cost	$/kWh-net	Case 10 Initial Cost	Case 10 Annual Cost	$/kWh-net	Case 10 Initial Cost	Case 10 Annual Cost	$/kWh-net	BN Initial Cost	BN Annual Cost	$/kWh-net	NOTE
162	B.2 Total Variable Operating Costs		**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	**Initial Cost**	**Annual Cost**	$/kWh-net	The annual cost is calculated on a 85% operating capacity (see p.52).
163	· Maintenance Material Cost		$ -	$ 8,763,218	0.00214	$ -	$ 9,308,185	0.00227	$ -	$ 15,644,314	0.00382	$ -	$16,617,203	0.00406	$ -	$ 43,925,932.86	0.01073	Scaling (Main. Mat Cost/TPC)$_{case\ 9,10}$ ~0.7%
164	· Consumables																	
165	- Water		$ -	$ 1,424,619	0.00035	$ -	$ 1,513,213	0.00037	$ -	$ 2,711,996	0.00066	$ -	$ 2,880,650	0.00070	$ -	$ 2,781,165	0.00068	Use DOE's price (p.379) ~ 0.00115 $/US gallon
	Daily water consumption @ 100% CF	(1000 gal/D)		4,245			4,245			8,081			8,081			7,802		
	Annual water consumption @ 85% CF	(1000 gal/year)		1,317,048			1,317,048			2,507,130			2,507,130			2,420,545.28		
	Cost of Water	($/1000 gal)		$ 1.08			$ 1.15			$ 1.08			$ 1.15			$ 1.15		
166	- Chemicals																	
167	MU & WT Chem.		$ -	$ 1,103,371	0.00027	$ -	$ 1,171,987	0.00029	$ -	$ 2,100,447	0.00051	$ -	$ 2,231,070	0.00054	$ -	$ 5,897,612.00	0.00144	Scaling (MU&WT/TPC)$_{case\ 9,10}$ ~0.1%
168	Limestone (FGD)		$ -	$ 3,496,290	0.00085	$ -	$ 3,713,717	0.00091	$ -	$ 5,043,346	0.00123	$ -	$ 5,356,982	0.00131	$ -	$ 6,071,581	0.00148	Scaling coal flow rate linearly (Line: 54)
169	Carbon (Mercury Removal)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	
170	MEA Solvent (MEA Carbon Capture)		$ -	$ -	-	$ -	$ -	-	$ 2,513,263	$ 1,105,563	0.00027	$ 2,669,558	$ 1,174,316	0.00029	$ -	$ -	-	
171	NaOH (SO_2 Polishing Scrubber - MEA)		$ -	$ -	-	$ -	$ -	-	$ 34,221	$ 1,061,704	0.00026	$ 36,349	$ 1,127,729	0.00028	$ 41,198	$ 2,556,327	0.00062	Scaling coal flow rate (Case 10) linearly (x2 to get to 1 ppm)
172	H_2SO_4		$ -	$ -	-	$ -	$ -	-	$ 10,450	$ 324,217	0.00008	$ 11,100	$ 344,379	0.00008	$ 12,581	$ 390,318	0.00010	Scaling coal flow rate linearly (Line: 54)

Calculation Sheet

Customer	ADA-ES	**Proj No**	111100208
Project Title	Solid Sorbent CO_2 Capture Pilot System	**Calc No**	
Calculation Title	Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant	**Phase/CTR**	
Elec File Location			
Project File Location	V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx		

INITIAL DESIGN / DOE INDIRECT COSTS

Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked
Elec File Location	26-Nov-15	BT	MER																				

Line No.	Description	unit	CASE 9 - (REF. PP)						CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)						CASE 10 WITH POLYSTYRENE BN SORBENT			NOTE
173	Corrosion Inhibitor (MEA Carbon Capture)		$ -	$ -	-	$ -	$ -	-	$ 154,511	$ 7,358	0.00000	$ 164,120	$ 7,816	0.00000	$ -	$ -	-	
174	Activated Carbon (MEA Carbon Capture)		$ -	$ -	-	$ -	$ -	-	$ -	$ 616,433	0.00015	$ -	$ 654,768	0.00016	$ -	$ -	-	
175	Ammonia (19% NH_3) (SCR)		$ -	$ 3,136,289	0.00077	$ -	$ 3,331,329	0.00081	$ -	$ 4,446,378	0.00109	$ -	$ 4,722,889	0.00115	$ -	$ 5,352,903	0.00131	Scaling coal flow rate linearly (Line: 54)
176	Solid Sorbent		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ 168,183,335	$ 51,746,077	0.01264	¥ % attrition rate (0.005%) and CO_2 removal rate
	Initial fill														Total mass, lb (44 trains)			
	Adsorber-1														169,450			
	Adsorber-2														134,562			
	Adsorber-3														383,401			
	Regenerator														69,700			
	Total	(lbm)													757,112			
	Total for 44 Trains	(lbm)													33,312,929			
	Grand Total	(Ton)													**16,656**			
	Cost of Sorbent	($/ton)													$ 10,097			
	Solid Comsumption (Mainly Attrition)																	
	%Attrition	(% / Cycle)														0.005		
	Total Sorbent flow rate	(lbm/hr)														27,530,510		The solid circ. rate linearly depends on the CO_2 captured flow rate.
	Spent solid	(lbm/hr)														1,377		
	Daily solid consumption @ 100% CF	(ton/day)														16.52		
	Annual solid consumption @ 85% CF	(ton/year)														5,125		
177	NaOH (CT - Solid CC)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 204,717	0.00005	Scaling CW requirement for CO_2 capture (83)
	Old CW recir. Rate	(gpm)														27,300		
	Old NaOH requirement	(lbm/day)														300.30		
	New NaOH requirement	(lbm/day)														2,864.85		
	Daily NaOH consumption @ 100% CF	(ton/day)														1.43		
	Annual NaOH consumption @ 85% CF	(ton/year)														444.41		
	Cost of NaOH	($/ton)														460.65		
178	Lime (CT - Solid CC)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 646,170	0.00016	Scaling CW requirement for CO_2 capture (83)
	Old CW recir. Rate	(gpm)														27,300		
	Old Lime requirement	(lbm/day)														1,965.60		
	New Lime requirement	(lbm/day)														18,751.76		
	Daily Lime consumption @ 100% CF	(ton/day)														9.38		
	Annual Lime consumption @ 85% CF	(ton/year)														2,908.87		
	Cost of Lime	($/ton)														222.14		
179	H_2SO_4 (CT - Solid CC)		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 220,354	0.00005	Scaling CW requirement for CO_2 capture (83)
	Old CW recir. Rate	(gpm)														27,300		
	Old H_2SO_4 requirement	(lbm/day)														1,010.10		
	New H_2SO_4 requirement	(lbm/day)														9,636.32		
	Daily H_2SO_4 consumption @ 100% CF	(ton/day)														4.82		
	Annual H_2SO_4 consumption @ 85% CF	(ton/year)														1,494.83		
	Cost of H_2SO_4	($/ton)														147.41		
180	**Subtotal Chemicals**		**$ -**	**$ 7,735,950**	0.00189	**$ -**	**$ 8,217,033**	0.00201	**$ 2,712,445**	**$ 14,705,446**	0.00359	**$ 2,881,127**	**$15,619,949**	0.00381	**$ 168,237,113**	**$ 73,086,061**	0.01785	
181	- Other																	
182	Supplemental Fuel		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	
183	SCR Catalyst		$ -	$ 592,641	0.00014	$ -	$ 629,496	0.00015	$ -	$ 831,516	0.00020	$ -	$ 883,226	0.00022	$ -	$ 1,001,045	0.00024	Scaling coal flow rate linearly (Line: 54)
184	Emission Penalties		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	
185	**Subtotal Other**		**$ -**	**$ 592,641**	0.00014	**$ -**	**$ 629,496**	0.00015	**$ -**	**$ 831,516**	0.00020	**$ -**	**$ 883,226**	0.00022	**$ -**	**$ 1,001,045**	0.00024	
186	- Waste Disposal																	
187	Fly Ash		$ -	$ 2,049,540	0.00050	$ -	$ 2,176,997	0.00053	$ -	$ 2,881,846	0.00070	$ -	$ 3,061,062	0.00075	$ -	$ 3,469,395	0.00085	Scaling coal flow rate linearly (Line: 54)
188	Bottom Ash		$ -	$ 512,385	0.00013	$ -	$ 544,249	0.00013	$ -	$ 720,462	0.00018	$ -	$ 765,266	0.00019	$ -	$ 867,349	0.00021	Scaling coal flow rate linearly (Line: 54)
189	Solid Waste		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ 88,348	0.00002	Assume 0.005% Attrition per cycle for 27,530,510 lbm solid/hr
	Solid Waste equal to Solid Consumption																	
	Cost of Solid Waste Treament	($/ton)														$ 17.24		
190	Waste water		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	Assume wastewater cost is accounted in the chemical costs.
191	**Subtotal Waste Disposal**		**$ -**	**$ 2,561,925**	0.00063	**$ -**	**$ 2,721,246**	0.00066	**$ -**	**$ 3,602,308**	0.00088	**$ -**	**$ 3,826,328**	0.00093	**$ -**	**$ 4,425,093**	0.00108	
192	- By-products & Emissions																	
193	Gypsum		$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	$ -	$ -	-	Cost of Gypsum Byproduct is zero.
194	**Subtotal By-Product**		**$ -**	**$ -**	-	**$ -**	**$ -**	-	**$ -**	**$ -**	-	**$ -**	**$ -**	-	**$ -**	**$ -**	-	
195	**Grand Total Variable Operating Costs**		**$ -**	**$21,078,353**	0.00515	**$ -**	**$22,389,174**	0.00547	**$ 2,712,445**	**$ 37,495,580**	0.00916	**$ 2,881,127**	**$39,827,356**	0.00973	**$ 168,237,113**	**$ 125,219,297**	0.03058	
196																		
197	B.3 Fuel (Variable Operating Costs)																	
198	· Coal (for Power Plant)		$ -	$62,175,575	0.01518	$ -	$66,042,151	0.01613	$ -	$ 87,425,787	0.02135	$ -	$92,862,624	0.02268	$ -	$ 105,250,115	0.02570	Calculated from coal feed rate with DOE's cost
	Coal flow rate	(lbm/hr)		437,378			437,378			614,994			614,994			697,032		

Calculation Sheet

Customer ADA-ES
Project Title Solid Sorbent CO_2 Capture Pilot System
Calculation Title Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant
Elec File Location
Project File Location V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx

INITIAL DESIGN / DOE INDIRECT COSTS

Proj No 111100208
Calc No
Phase/CTR

Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked
Elec File Location	26-Nov-15	BT	MER																

Line No.	Description	unit	CASE 9 - (REF. PP)						CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)						CASE 10 WITH POLYSTYRENE BN SORBENT			NOTE
	Daily coal consumption @ 100% CF	(ton/day)		5,249			5,249			7,380			7,380			8,364		
	Annual coal consumption @ 85% CF	(ton/year)		1,628,358			1,628,358			2,289,624			2,289,624			2,595,050		
	Cost of coal	($/ton)		$ 38.18			$ 40.56			$ 38.18			$ 40.56			$ 40.56		
199	**Total Fuel Costs**		**$ -**	**$62,175,575**	0.01518	**$ -**	**$66,042,151**	0.01613	**$ -**	**$ 87,425,787**	0.02135	**$ -**	**$92,862,624**	0.02268	**$ -**	**$ 105,250,115**	0.02570	
200																		
201	**C. CO_2 Transport, Storage & Monitoring**																	
202	C.1 Transport Costs		June		2007	Feb		2011	June		2007	Feb		2011	Feb		2011	
203	· Pipeline Costs		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
204	· Related Capital Expenditures		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
205	· O&M Costs		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
206	C.2 Storage Costs																	
207	· Site Screening and Evaluation		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
208	· Injection Wells		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
209	· Injection Equipment		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
210	· O&M Costs		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
211	· Pore Volume Acquisition		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
212	C.3 Monitoring Costs		$	-	-	$	-	-	$	-	-	$	-	-	$	-	-	
213	**Total CO_2 TS&M Costs**		**$**	**-**	-	**$**	**-**	-	**$**	**23,751,018**	43	**$**	**25,228,047**	46	**$**	**28,543,657**	52	Scaling CO_2 Removal Rate line((Line: 63)
214																		
215	**Economic Parameters**		June		2007	Feb		2011	June		2007	Feb		2011	Feb		2011	
216	Type of Risk		Low			Low			High			High			High			
217			COE	LCOE		COE	LCOE		COE	LCOE		COE	LCOE		COE	LCOE		
218	1. Capital Charge Factor, 5 Years		0.116	0.116		0.116	0.116		0.124	0.124		0.124	0.124		0.124	0.124		
219	2. Nominal (Apparent) Escalation Rate		3%	0%		3%	0%		3%	0%		3%	0%		3%	0%		
220	3. General Inflation Rate		3%	0%		3%	0%		0%	0%		0%	0%		0%	0%		
221	4. Real Escalation Rate		0%	0%		0%	0%		0%	0%		0%	0%		0%	0%		
222	5. Debt (% of Total)		50	50		50	50		45	45		45	45		45	45		
223	6. Equity (% of Total)		50	50		50	50		55	55		55	55		55	55		
224	7. Real Discount Rate (Nominal)		8.25%	8.25%		8.25%	8.25%		9.08%	9.08%		9.08%	9.08%		9.08%	9.08%		
225	8. Real Discount Rate (After Tax)		7.39%	7.39%		7.39%	7.39%		8.13%	8.13%		8.13%	8.13%		8.13%	8.13%		
226																		
227	**COE contribution (¢/kWh)**		June		2007	Feb		2011	June		2007	Feb		2011	Feb		2011	**PROCEDURE 2 - LCOE Calculation**
228	1. Capital Costs		3.11			3.30			6.01			6.39			16.88			LCOE is obtained when the difference between the NPV of COE and
229	2. Fixed Operating Costs		0.78			0.83			1.31			1.39			3.21			LCOE is equal to zero.
230	3. Variable Operating Costs		0.51			0.55			0.92			0.97			3.06			Step 1: Change cost parameters (Lines 123 - 230).
231	4. Fuel Costs		1.52			1.61			2.13			2.27			2.57			Step 2: Use Solver in Data Tab. In the Solver's windor, set parameters
232	5. CO_2 T&SM Costs		0.00			0.00			0.58			0.62			0.70			as follows:
233	**Total COE (2007)**		**5.93**	**7.53**		**6.29**	**8.00**		**10.95**	**13.9**		**11.63**	**14.77**		**26.42**	**33.49**		- Set Target Cell: R400
234			**5.94**			**6.31**			**10.97**			**11.65**						- Select option "Value of" and set the blank cell equal to 0.
235	**Total COE (2012, @ 3% average annual inflation rate)**		6.89	8.73		7.31	9.27		12.72	16.1		13.51	17.12		30.62	38.82		- By Changing Cells: T358
236																		- Click Solve button to get new LCOE.

Line No.	Year			COE	LCOE	COE	LCOE	COE	LCOE	COE	LCOE	COE	LCOE	NOTE
237	**Year**			**COE**	**LCOE**	**COE**	**LCOE**	**COE**	**LCOE**	**COE**	**LCOE**	**COE**	**LCOE**	
238	**2007**	**0**	ase Year - Con	(2.681)	(2.681)	(2.848)	(2.848)	(4.848)	(4.848)	(5.149)	(5.149)	(13.61)	(13.61)	
239	**2008**	**1**	Construction	(8.334)	(8.334)	(8.852)	(8.852)	(15.067)	(15.067)	(16.004)	(16.004)	(42.31)	(42.31)	
240	**2009**	**2**	Construction	(7.195)	(7.195)	(7.642)	(7.642)	(13.008)	(13.008)	(13.817)	(13.817)	(36.52)	(36.52)	
241	**2010**	**3**	Construction	(5.963)	(5.963)	(6.334)	(6.334)	(10.781)	(10.781)	(11.452)	(11.452)	(30.27)	(30.27)	
242	**2011**	**4**	Construction	(4.633)	(4.633)	(4.921)	(4.921)	(8.377)	(8.377)	(8.898)	(8.898)	(23.52)	(23.52)	
243	**2012**	**5**	Operating	6.886	8.728	7.314	9.271	12.72	16.120	13.51	17.122	30.62	38.820	
244	**2013**	**6**	Operating	7.093	8.728	7.534	9.271	13.10	16.120	13.91	17.122	31.54	38.820	
245	**2014**	**7**	Operating	7.305	8.728	7.760	9.271	13.49	16.120	14.33	17.122	32.49	38.820	
246	**2015**	**8**	Operating	7.525	8.728	7.993	9.271	13.90	16.120	14.76	17.122	33.46	38.820	
247	**2016**	**9**	Operating	7.750	8.728	8.232	9.271	14.31	16.120	15.20	17.122	34.47	38.820	
248	**2017**	**10**	Operating	7.983	8.728	8.479	9.271	14.74	16.120	15.66	17.122	35.50	38.820	
249	**2018**	**11**	Operating	8.222	8.728	8.734	9.271	15.18	16.120	16.13	17.122	36.57	38.820	
250	**2019**	**12**	Operating	8.469	8.728	8.996	9.271	15.64	16.120	16.61	17.122	37.66	38.820	
251	**2020**	**13**	Operating	8.723	8.728	9.266	9.271	16.11	16.120	17.11	17.122	38.79	38.820	
252	**2021**	**14**	Operating	8.985	8.728	9.544	9.271	16.59	16.120	17.62	17.122	39.96	38.820	
253	**2022**	**15**	Operating	9.254	8.728	9.830	9.271	17.09	16.120	18.15	17.122	41.16	38.820	**NOTE**
254	**2023**	**16**	Operating	9.532	8.728	10.125	9.271	17.60	16.120	18.70	17.122	42.39	38.820	**1. If the performance parameters and economic parameters are**
255	**2024**	**17**	Operating	9.818	8.728	10.428	9.271	18.13	16.120	19.26	17.122	43.66	38.820	**changed, follow Procedures 1 and 2.**
256	**2025**	**18**	Operating	10.112	8.728	10.741	9.271	18.67	16.120	19.84	17.122	44.97	38.820	**2. If only the performance parameters are changed, follow Procedures**
257	**2026**	**19**	Operating	10.416	8.728	11.064	9.271	19.23	16.120	20.43	17.122	46.32	38.820	**1 and 2.**

Calculation Sheet

Customer ADA-ES
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Calculation Title Performance and Cost Analysis for an Integration of Sorbent BN-based CO_2 Capture with a 550 MW (net) Subcritical Pulverized Coal-fired Power Plant
Elec File Location
Project File Location V:\1111\active\111100080\design\1900\230\lcoe_case10_ADA_bt_20110530_rev.E.xlsx

INITIAL DESIGN / DOE INDIRECT COSTS

Proj No 111100208
Calc No
Phase/CTR

Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked	Rev	Date	By	Checked
Elec File Location	26-Nov-15	BT	MER																

Line No.	Description			unit	CASE 9 - (REF. PP)						CASE 10 - (REF. PP WITH FLUOR'S AMINE BASED CAPTURE)						CASE 10 WITH POLYSTYRENE BN SORBENT				NOTE
258	**2027**	**20**	Operating		10.728	8.728		11.395	9.271		19.81	16.120		21.04	17.122			47.71	38.820		**3. If only the economic parameters are changed, follow Procedure 2.**
259	**2028**	**21**	Operating		11.050	8.728		11.737	9.271		20.41	16.120		21.67	17.122			49.14	38.820		**4. A change in Thermal Input affects the CO_2 Captured Flow Rate and**
260	**2029**	**22**	Operating		11.382	8.728		12.089	9.271		21.02	16.120		22.32	17.122			50.62	38.820		**Gross Output.**
261	**2030**	**23**	Operating		11.723	8.728		12.452	9.271		21.65	16.120		22.99	17.122			52.13	38.820		
262	**2031**	**24**	Operating		12.075	8.728		12.826	9.271		22.30	16.120		23.68	17.122			53.70	38.820		
263	**2032**	**25**	Operating		12.437	8.728		13.210	9.271		22.97	16.120		24.39	17.122			55.31	38.820		
264	**2033**	**26**	Operating		12.810	8.728		13.607	9.271		23.66	16.120		25.13	17.122			56.97	38.820		
265	**2034**	**27**	Operating		13.194	8.728		14.015	9.271		24.37	16.120		25.88	17.122			58.68	38.820		
266	**2035**	**28**	Operating		13.590	8.728		14.435	9.271		25.10	16.120		26.66	17.122			60.44	38.820		
267	**2036**	**29**	Operating		13.998	8.728		14.869	9.271		25.85	16.120		27.46	17.122			62.25	38.820		
268	**2037**	**30**	Operating		14.418	8.728		15.315	9.271		26.62	16.120		28.28	17.122			64.12	38.820		
269	**2038**	**31**	Operating		14.850	8.728		15.774	9.271		27.42	16.120		29.13	17.122			66.04	38.820		
270	**2039**	**32**	Operating		15.296	8.728		16.247	9.271		28.25	16.120		30.00	17.122			68.02	38.820		
271	**2040**	**33**	Operating		15.755	8.728		16.735	9.271		29.09	16.120		30.90	17.122			70.06	38.820		
272	**2041**	**34**	Operating		16.228	8.728		17.237	9.271		29.97	16.120		31.83	17.122			72.17	38.820		
273	**2042**	**35**	Operating		16.714	8.728		17.754	9.271		30.87	16.120		32.78	17.122			74.33	38.820		
274				**NPV**	**18.43**	**18.43**		**19.57**	**19.57**		**34.89**	**34.89**		**37.06**	**37.06**			**69.26**	**69.26**		
275				**D NPV**	**(0.00)**			**(0.00)**			**(0.00)**			**(0.00)**				-			
276				**IRROE**		**0.1225**			**0.1225**			**0.1224**			**0.1224**				**0.1224**		