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Bench-Scale Process for Low-Cost Carbon Dioxide (CO_2) Capture Using a Phase-Changing Absorbent

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Principal Investigator:	Tiffany Westendorf Chemical Engineer 518-387-7347 westendo@research.ge.com
Contributors:	Dwayne McDuffie, Barbara Miebach, Irina Spiry
Recipient Organization:	GE Global Research 1 Research Circle Niskayuna, NY 12309-1027
Submitted To:	U.S. Department of Energy National Energy Technology Laboratory 626 Cochrans Mill Road P.O. Box 10940 Pittsburgh, PA 15236-0940
Signature of Submitting Official	Tiffany Westendorf

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Abstract

The objective of this project is to design and build a bench-scale process for a novel phase-changing CO_2 capture solvent. The project will establish scalability and technical and economic feasibility of using a phase-changing CO_2 capture absorbent for post-combustion capture of CO_2 from coal-fired power plants with 90% capture efficiency and 95% CO_2 purity at a cost of \$40/tonne of CO_2 captured by 2025 and a cost of <\$10/tonne of CO_2 captured by 2035. This report presents system and economic analysis for a process that uses a phase changing aminosilicone solvent to remove CO_2 from pulverized coal (PC) power plant flue gas. The aminosilicone solvent is a pure 1,3-bis(3-aminopropyl)-1,1,3,3-tetramethyldisiloxane (GAP-0). Performance of the phase-changing aminosilicone technology is compared to that of a conventional carbon capture system using aqueous monoethanolamine (MEA).

This analysis demonstrates that the aminosilicone process has significant advantages relative to an MEA-based system. The first-year CO_2 removal cost for the phase-changing CO_2 capture process is \$52.1/tonne, compared to \$66.4/tonne for the aqueous amine process. The phase-changing CO_2 capture process is less costly than MEA because of advantageous solvent properties that include higher working capacity, lower corrosivity, lower vapor pressure, and lower heat capacity. The phase-changing aminosilicone process has approximately 32% lower equipment capital cost compared to that of the aqueous amine process. However, this solvent is susceptible to thermal degradation at CSTR desorber operating temperatures, which could add as much as \$88/tonne to the CO_2 capture cost associated with solvent makeup. Future work is focused on mitigating this critical risk by developing an advanced low-temperature desorber that can deliver comparable desorption performance and significantly reduced thermal degradation rate.

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Executive Summary

The objective of this project is to design and build a bench-scale process for a novel phase-changing CO_2 capture solvent. The project will establish scalability and technical and economic feasibility of using a phase-changing CO_2 capture absorbent for post-combustion capture of CO_2 from coal-fired power plants with 90% capture efficiency and 95% CO_2 purity at a cost of \$40/tonne of CO_2 captured by 2025 and a cost of <\$10/tonne of CO_2 captured by 2035.

This report presents system and economic analysis for a process that uses a phase changing aminosilicone solvent to remove CO_2 from pulverized coal (PC) power plant flue gas. The aminosilicone solvent is a pure 1,3-bis(3-aminopropyl)-1,1,3,3-tetramethyldisiloxane (GAP-0). Performance of the phase-changing aminosilicone technology is compared to that of a conventional carbon capture system using aqueous monoethanolamine (MEA).

The phase-changing CO_2 capture system is a thermal-swing absorption/desorption process that utilizes a spray absorber and stirred-tank (CSTR) desorber. The solvent absorbs CO_2 from power plant flue gas, forming a solid. This solid mixes with water that is present in the flue gas and contained in the solvent to form a pumpable, CO_2-rich slurry. This slurry is pumped to a CSTR desorber, wherein CO_2 is produced by heating the slurry to reverse the chemical reaction. Regenerated solvent is recycled to the absorber for reuse. Process utilities needed to operate the CO_2 capture system, such as power and steam, are supplied by the power plant.

Aspen Plus models were developed for both the MEA and phase-changing aminosilicone CO_2 capture processes to calculate the mass and energy balances and system performance. The models account for steam load for the CO_2 capture systems and parasitic loads for solvent pumps, CO_2 compressors, and cooling water pumps. The power plant was modeled using Thermoflow. Cost and performance estimates were developed for a $550MW_{net}$ coal-fired power plant with CO_2 capture, and reflect heat integration between the power plant and the CO_2 capture system to maximize overall efficiency.

This analysis demonstrates that the aminosilicone process has significant advantages relative to an MEA-based system. The first-year CO_2 removal cost for the phase-changing CO_2 capture process is \$52.1/tonne, compared to \$66.4/tonne for the aqueous amine process. The phase-changing CO_2 capture process is less costly than MEA because of advantageous solvent properties that include higher working capacity, lower corrosivity, lower vapor pressure, and lower heat capacity. Higher working capacity reduces the solvent flow rate required for 90% capture, resulting in smaller equipment. Lower solvent corrosivity allows for selection of less costly materials of construction (i.e., carbon steel) for key unit operations, including the rich/lean heat exchanger. Lower vapor pressure results in reduced solvent losses to the gas product streams. Lower heat capacity increases the energy efficiency of the process by decreasing the sensible heat duty in the desorber and rich lean heat exchanger. These advantages result in approximately 32% lower equipment capital cost of the phase-changing CO_2 capture process compared to that of the aqueous amine process.
However, the aminosilicone solvent is susceptible to thermal degradation at CSTR desorber operating temperatures, which could add as much as \$88/tonne to the CO_2 capture cost

associated with solvent makeup. Future work is focused on mitigating this critical risk by developing an advanced low-temperature desorber that can deliver comparable desorption performance and significantly reduced thermal degradation rate.

Completion of Task 7.2: Techno-economic assessment

The project will establish scalability and technical and economic feasibility of using a phase-changing CO_2 capture absorbent for post-combustion capture of CO_2 from coal-fired power plants with 90% capture efficiency and 95% CO_2 purity at a cost of $40/tonne of CO_2 captured by 2025 and a cost of <$10/tonne of CO_2 captured by 2035. This 45-month project is divided into three budget periods with milestones and two Go/No-Go decisions. At the close of the project, ending 9/30/2017, the cost of CO_2 capture from a $550MW_{net}$ coal-fired power plant should be less than $40/tonne of CO_2 captured, assuming 90% CO_2 capture and 95% CO_2 purity. This report discusses details of the analysis leading to the cost estimates for the phase-changing CO_2 capture process utilizing a CSTR desorber.

In a previous effort[1], a power plant model without carbon capture was developed to mimic Case 11 of the DOE Bituminous Baseline Study (BBS)[2]. That model was used to support techno-economic analysis of the phase-changing aminosilicone process. Cost and performance estimates of the phase-changing aminosilicone CO_2 capture process developed in this project were compared to Case 12 of the BBS, which utilizes an aqueous amine solvent.

Process Description

A supercritical pulverized coal (PC) plant and CO_2 separation unit based on monoethanolamine (MEA) is described in Case 12 of the DOE BBS[2]. This case is the benchmark against which the phase-changing CO_2 capture process will be compared. A simplified block diagram of the power plant and CO_2 separation system is shown in Figure 1. The pulverized coal boiler generates steam, which is sent to the steam turbines. The flue gas is sent through a selective catalytic reduction (SCR) unit to reduce nitrogen oxides (NO_X), a bag house to remove fly ash, and a flue gas desulfurizer (FGD) to remove sulfur dioxide (SO_2). The flue gas is then sent through the CO_2 separation unit before being vented to the stack.

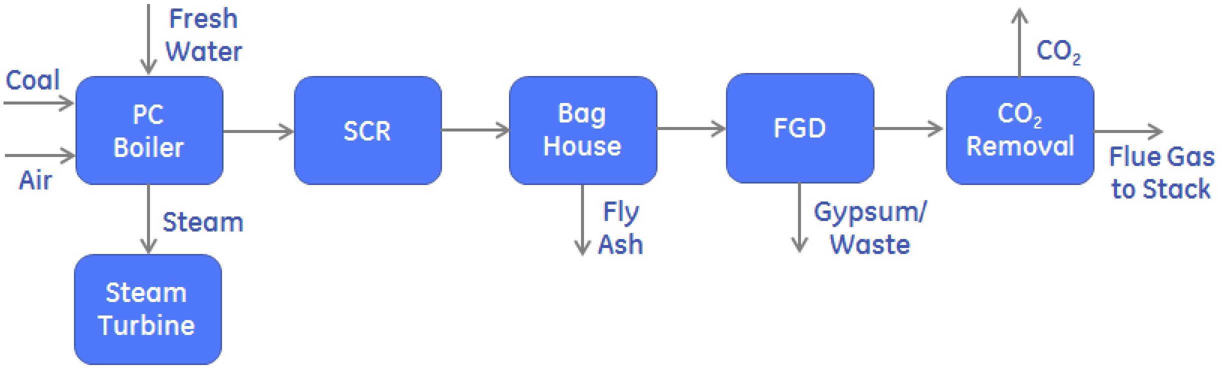


Figure 1. Process flow diagram of coal-fired power plant with CO_2 capture.

Power Plant Modeling in Thermoflow

A model of a supercritical pulverized coal plant was built in Thermoflow, a thermodynamic design tool which includes cost estimation methods for conventional coal-fired power plants. The Thermoflow model interacted with the carbon capture model by exchanging flue gas, process steam, and water at the boundaries between the two systems. Capital costs, operating costs, and net power output were rolled up at the plant level. The pulverized coal power plant model was developed under a previous project[1] to mimic Case 11 of the DOE BBS[2] (DOE Case 11).

In the previous study, the Thermoflow power plant model was calibrated to DOE Case 11. Gas and steam flows, pressures and temperatures throughout the plant, exhaust composition, auxiliary loads, and net plant output predicted by the Thermoflow model were closely matched to DOE Case 11. Efficiencies of all the major equipment in the power block, including pumps, fans, steam turbine sections, the boiler, and environmental equipment were specified to match DOE Case 11, and were fixed as the model was updated to include CO_2 capture. The cost model in Thermoflow was also tuned to match the overall capital costs given in DOE Case 11. Because the cost breakdown in Thermoflow's cost estimation tool is not as detailed as that in the BBS, the plant capital costs were calibrated only for the full plant rather than on a component level. This power plant model was used to develop the cost estimates for the phase-changing aminosilicone CO_2 capture process, as discussed in a later section. Figure 2 shows a simplified block diagram of the power block for this mode. For detailed process flow information for each stream in Figure 2, see Tables 1-6 in the Preliminary Process and Cost Modeling report[3].

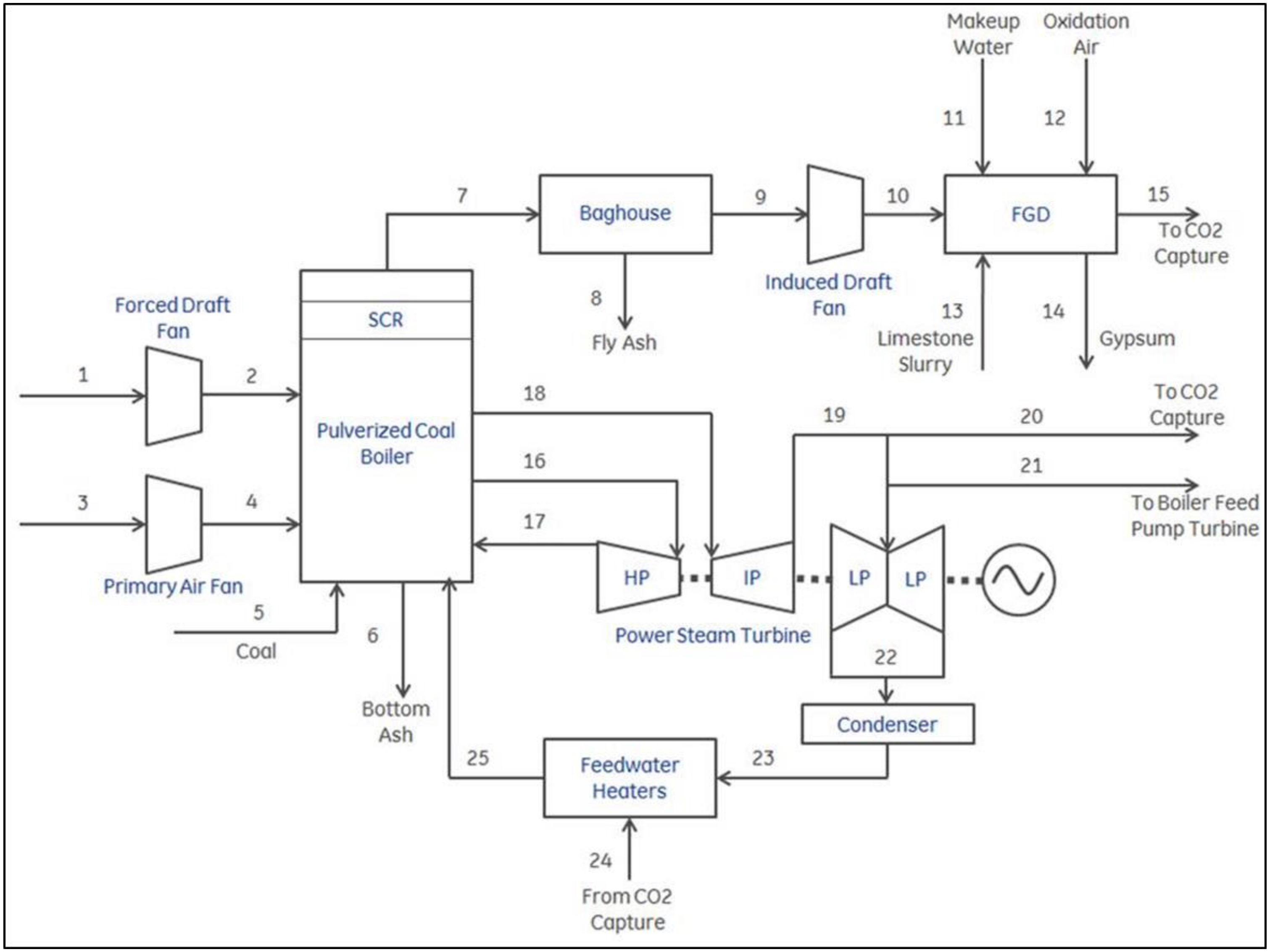


Figure 2. Block flow diagram for power plant without CO_2 capture (comparable to DOE Case 11).

CO_2 Capture System Aspen Plus Model Development

The MEA and GAP-0 CO_2 separation units utilize four key processes, CO_2 absorption, CO_2 desorption, solvent handling, and CO_2 compression. The flue gas from the power plant is processed in a direct contact cooler to reduce the temperature to 40°C (104°F) and then enters the absorber. The direct contact cooler also acts as a polishing sulfur scrubber, as discussed in a later section of this report.

An Aspen Plus model was developed for a commercial-scale CO_2 capture process using phase-changing aminosilicone solvent. CO_2 capture process models were developed for several different sensitivity cases, varying absorber and desorber operating conditions and configuration. For all these cases, the flue gas flow rate was fixed to match that of DOE Case 11, which produces 550 MW net power without CO_2 capture. Comparison of these cases allowed for selection of a case that yielded the lowest overall cost of CO_2 removal. The best case was then scaled up to 550 MW net power with CO_2 capture. Detailed heat integration was carried out starting from the scaled-up best case to optimize the power plant with carbon capture. These cases are summarized in Table 1. The details of the selected cases are explained in the following sections. A schematic sketch of the phase-changing CO_2 capture process concept is shown in Figure 3.

Table 1. Summary of major cases considered for the phase-changing aminosilicone CO_2 separation system.

Case	# of absorber stages	Lean splits to absorb stages	Desorber Temperature (°C)	Rich/Lean U (BTU/hr/ft^2/F)	Rich Solvent	Rich pump	550MW net w/ capture
6F	10	10 equal	140	75	5 wt% H_2O	Slurry Pump	N
6J	3	3 equal	140	75	5 wt% H_2O	Slurry Pump	N
6K	3	3 equal	140	75	10 wt% H_2O	Slurry Pump	N
6L	3	3 equal	140	75	15 wt% H_2O	Slurry Pump	N
6M	3	50/25/25	140	75	5 wt% H_2O	Slurry Pump	N
6N	3	25/50/25	140	75	5 wt% H_2O	Slurry Pump	N
6O	3	25/25/50	140	75	5 wt% H_2O	Slurry Pump	N
6P	4	4 equal	140	75	5 wt% H_2O	Slurry Pump	N
6Q	3	3 equal	140	75	5 wt% H_2O	Slurry Pump	Y
6R	3	3 equal	140	75	5 wt% H_2O	Slurry Pump	Y

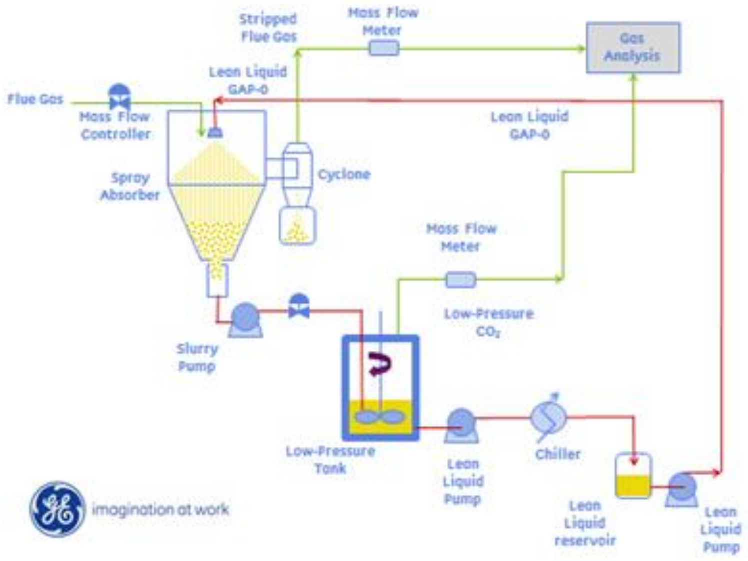


Figure 3. Conceptual design for the phase changing aminosilicone CO_2 capture process.

An overview of the phase-changing aminosilicone process model is shown in Figure 4 and in Figure 5. The lean solvent enters the absorber at 40°C (104°F) and chemically reacts with CO_2 contained in the flue gas, forming a slurry. The absorber for this process is a spray absorber, since the rich solvent leaves the absorber in a slurry form. The exothermic CO_2 absorption reaction increases the temperature of the solvent. The absorber is operated at 40-71°C (104-160°F) and at atmospheric pressure. The rich solvent from the absorber is fed to a slurry pump, which feeds the slurry into an atmospheric pressure desorber, wherein the solids are heated to 130-140°C (266 to 284°F) to produce a stream of CO_2 and a regenerated stream of solvent. The hot vapor products from the desorber, which consist primarily of CO_2, are cooled in heat exchangers utilizing water. Entrained solvent is removed from the cooled gas streams and returned to the desorber. The gas stream is further cooled to remove the entrained water, which is sent to the waste water treatment plant. The cooled CO_2 gas is delivered to the CO_2 product compressor. The lean solvent exiting the desorber is cooled before recycle to the absorber. In the commercial process, this cooling step would be part of a rich/lean heat exchange arrangement, to maximize heat management around the desorption system.

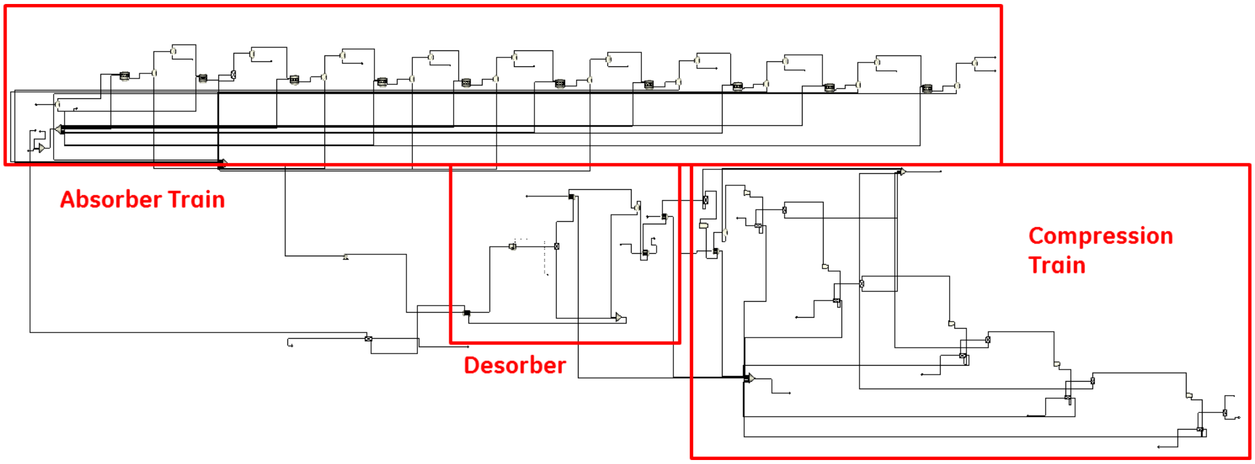


Figure 4. Aspen Plus model for CO_2 separation sub-system - Case 6F

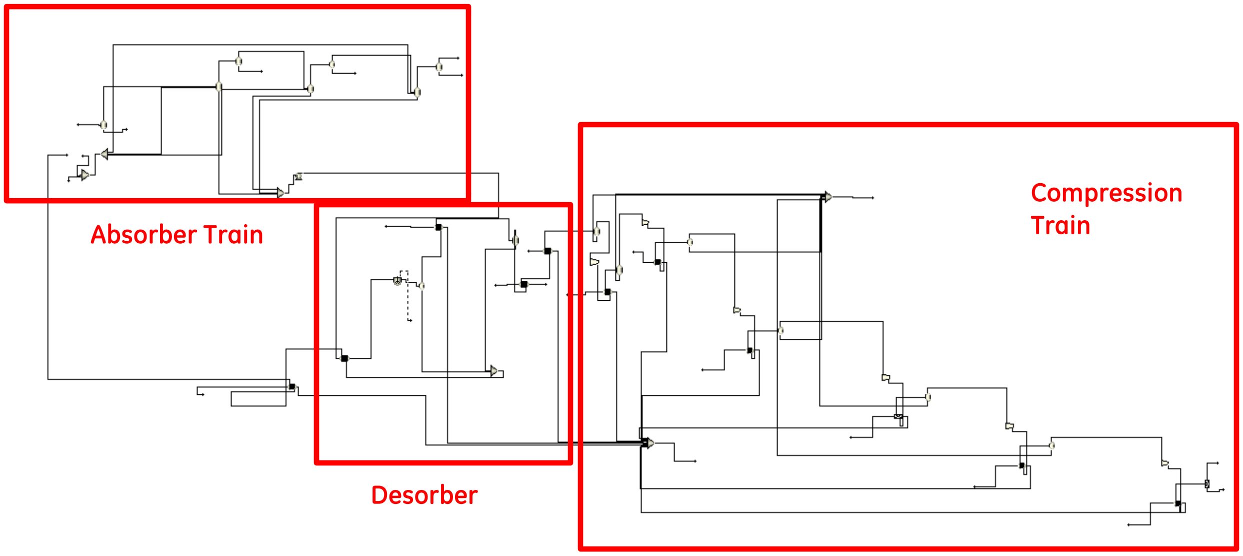


Figure 5. Aspen Plus model for CO_2 separation sub-system Cases 6J-6R

CO_2 Separation Unit Key Assumptions

The CO_2 separation process model used the following design assumptions given in DOE Case 11[2].

1) Composition of flue gas leaving the FGD (wet basis) is shown in Table 2.

Table 2. Flue gas composition leaving FGD.

	Volume %
CO_2	13.53
H_2O	15.17
N_2	68.9
O_2	2.40

SOx	0.25-42 ppmv
NOx	74 ppmv

2) The flow rate of flue gas leaving the FGD (based on DOE Case 11[2]): 4,713,221 lb/hr. The flow rate for the scaled-up cases varied due to differences in overall plant efficiency with the various CO_2 capture system configurations.
3) Pressure and temperature of flue gas leaving FGD: 14.8 psia and 135°F
4) Conditions for LP steam available from power plant: 556°F (base case, sensitivity was conducted with respect to steam conditions)
5) Conditions for cooling water: feed = 60°F, return = 80°F.
6) CO_2 removal from flue gas: greater than 90%
7) CO_2 purity: greater than 95 vol%
8) CO_2 delivery pressure and temperature: 2,215 psia and 124°F

The MEA and phase-changing aminosilicone solvent baseline models are based on a typical temperature-swing absorbent separation process. The systems have four process variables that dominate the performance with a given absorbent (solvent) and they are absorber temperature, desorber temperature, desorber pressure, and rich/lean heat exchanger approach temperature. The system models account for the major energy penalties for CO_2 separation, and they include the energy required:

(1) for vaporization of water
(2) to desorb the carbon dioxide (i.e., reaction energy)
(3) for sensible heating of the solvent

The energy is supplied by feeding steam to the desorber. The models also account for CO_2 compression energy and auxiliary loads.

The solvent rich loading is defined as the weight percent of CO_2 in the rich solvent leaving the absorber column. The solvent lean loading is defined as the weight percent of CO_2 in the lean solvent leaving the desorber column. The solvent net loading is defined as the difference between the rich loading and the lean loading and was obtained from experimental absorption and desorption isotherms for the GAP-0 system.

Absorber Design

Flue gas enters the CO_2 capture process from the coal-fired power plant. The flue gas flow rate and composition were determined from the results of the Thermoflow power plant model. The flue gas is cooled to 40°C in a direct contact cooler, where condensed water is removed and can be used as makeup water in the process or sent to a wastewater treatment plant. The flue gas is contacted with an atomized spray of the phase-changing aminosilicone CO_2 capture solvent in a multistage spray absorber. The phase-changing solvent solidifies upon reaction with CO_2, and mixes with water present in the flue gas and contained in the lean solvent to produce a pumpable carbamate/water slurry. The reaction is exothermic therefore the gas stream heats up as it flows through the absorber. Since the solvent absorbs CO_2 more efficiently at lower temperatures, the maximum temperature of each stage should be maintained below 80°C. To maintain low flue gas temperature, the flue gas stream is cooled to 40°C between stages by spraying the gas stream with water in direct contact coolers. The condensed water is removed and can be used as process makeup or sent to a wastewater treatment plant. The flue gas stream then flows to the next absorber stage for contact with the next spray stage of lean GAP-0 solvent.

In previous versions of the Aspen Plus model[3,7], the absorber unit operation was modeled as a series of stoichiometric reactors with specified fractional conversion of CO_2 and heat of reaction, as shown in Figure 4. This setup was used because insufficient experimental data was available to be incorporated into the absorber model. In 2016, the continuous bench scale system was built, and experimental data was generated that allowed for development of a more rigorous absorber model. In this more rigorous model, the absorber is modeled as an equilibrium flash reactor that uses a globally-defined equilibrium chemical reaction constant K_{eq} as shown in Figure 5. This approach allows for direct calculation of the absorber exit temperature and composition, whereas in previous versions of the model the absorber exit temperature was artificially limited by using a larger number of spray stages to achieve 90% capture overall. The impact of this change on the absorber design and cost is discussed in a later section of this report.

Bench scale experimental data was used to regress the parameters of this equilibrium reaction using the expression in Equation 1.

Equation 1. Chemical reaction equilibrium constant correlation.

$$lnK_{eq} = A + \frac{B}{T}$$

A randomized subset of the continuous bench scale absorber data set was segregated for regression validation. The rest of the data set was used to build the regression. The regression was built using MATLAB in conjunction with Aspen Plus. MATLAB automatically entered initial values for the equilibrium constant parameters A and B and absorber input parameters (defined in Table 3) into an absorber equilibrium flash reactor block in Aspen. MATLAB executed the Aspen simulation, calculated the residual error between the Aspen calculations and experimental data for rich solvent temperature and carbamate loading, and selected values for the equilibrium constant parameters to use in the next iteration. This iterative regression was repeated to minimize the residual error between the Aspen calculations and experimental data for rich solvent temperature and carbamate loading.

This regression yielded an Aspen absorber model that predicts within 20% of the experimentally observed rich solvent carbamate loading and temperature, as shown in Figure 6 and Figure 7. In Figure 7 it is apparent that the absorber model overpredicts rich solvent temperature compared to the experimental data. This discrepancy is attributed to heat losses that are present in the bench scale spray absorber, which are not accounted for in the Aspen Plus model. Overall, it was concluded that regression of equilibrium constant parameters yielded good agreement of the model with experimental data, and this absorber model was used to scale the phase-changing aminosilicone process up to $550MW_{net}$.

Table 3. Summary of input and output parameters used in MATLAB/Aspen regression of equilibrium constant parameters.

Stream	Parameter	Function
Lean Solvent In	Flow rate	Input into model, defined by experimental regression data set
	Temperature	Input into model, defined by experimental regression data set
	%Water	Input into model, defined by experimental regression data set
	%Carbamate	Input into model, defined by experimental regression data set
Gas In	Flow rate	Input into model, defined by experimental regression data set
	Temperature	Input into model, defined by experimental regression data set
	%CO_2	Input into model, defined by experimental regression data set
Gas Out	%CO_2	Output for comparison only
	Temperature	Output for comparison only
Rich Solvent Out	Flow Rate	Output for comparison only
	Temperature	Output for regression, algorithm minimizes the residual error of this result compared to experimental regression data set
	%Carbamate	Output for regression, algorithm minimizes the residual error of this result compared to experimental regression data set

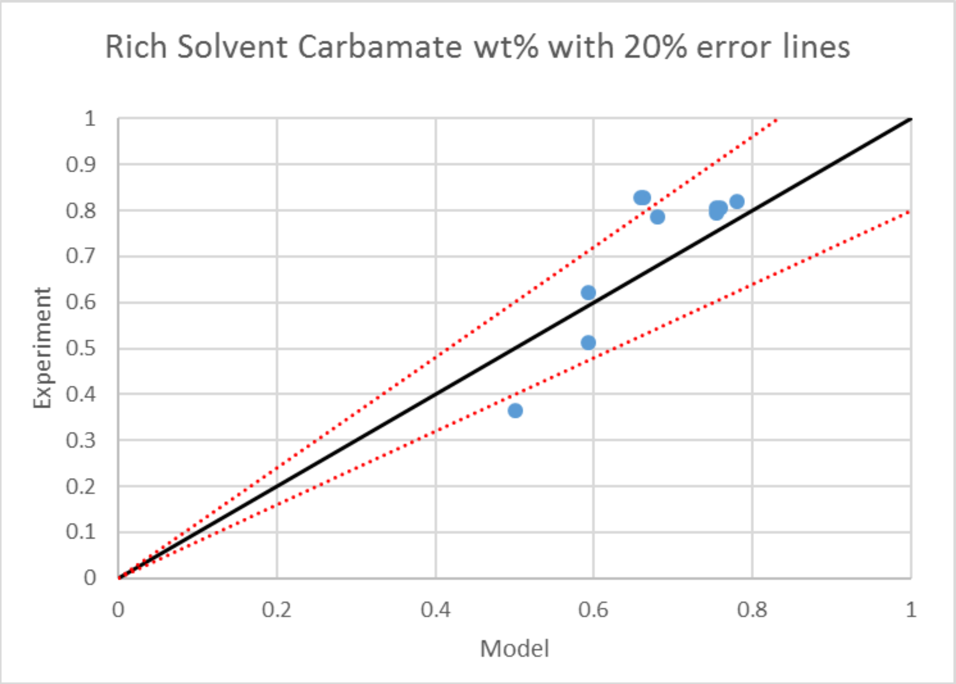


Figure 6. Comparison of rich solvent carbamate loading between experimental data and model prediction.

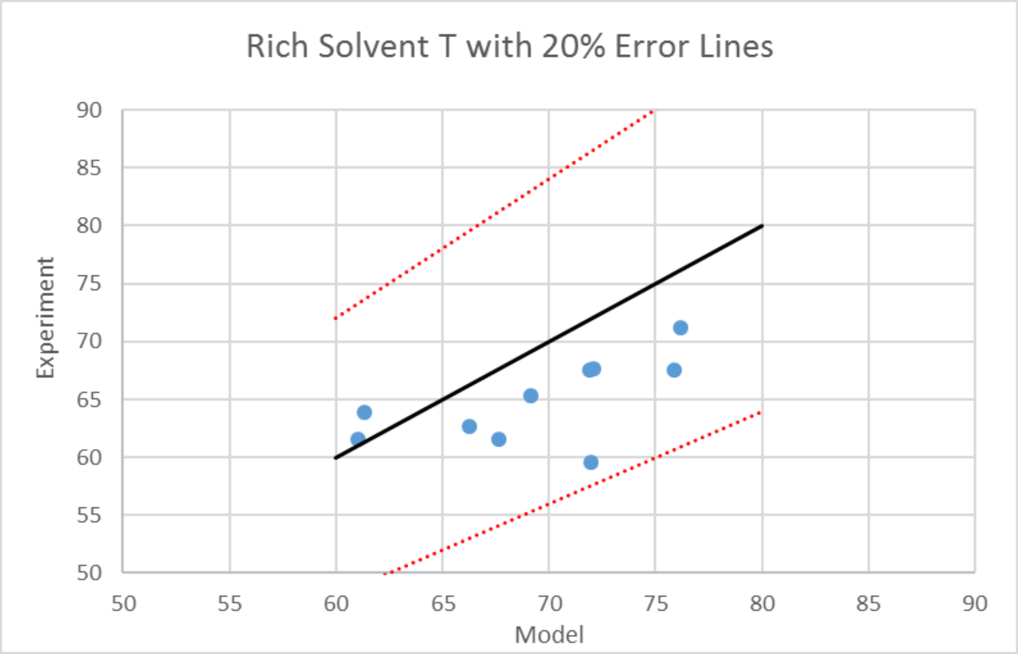


Figure 7. Comparison of rich solvent temperature between experimental data and model prediction.

Slurry Pump

The rich solvent stream is a slurry consisting of carbamate solids, water, and unreacted GAP-0. Current bench scale testing uses a progressive cavity pump to pump this slurry from the absorber to the desorber. Pumping power estimates for the techno-economic analysis were obtained from vendor quotes, based on knowledge of slurry viscosity at varying process conditions as predicted by Aspen.

A new correlation for viscosity as a function of slurry composition (GAP-0, carbamate, and water) and temperature was integrated into the Aspen Plus model. In previous versions of the model, viscosity of the carbamate was not included in the property set up. The challenge of including phase-changing solvent viscosity in the model was that pure GAP-0 carbamate is a solid, and therefore has an undefinable viscosity. In 2016, the phase-changing CO_2 capture process concept was modified to rely on a GAP-0 carbamate/water slurry as the rich solvent phase leaving the absorber. A thorough study of slurry viscosity was conducted[4], exploring the effects of temperature and composition. This data was regressed to be included in the Aspen Plus model to generate more rigorous estimates of slurry pumping power. The viscosity regression was performed with MATLAB and Aspen Plus using a similar approach to the Keq regression described above.

Experimental viscosity measurements were taken using a cup and bob viscometer. Multiple samples were prepared with compositions as presented in Table 1. Viscosity was measured for temperatures of 25-80°C and shear rates of 1-150s^{-1}. All viscosity data was found to be independent of shear rate, and the data for 25s^{-1} was used in the regression.

Table 4. Composition of samples for viscosity measurement.

Sample	***Water (wt%)***	***GAP-0 (wt%)***	***Carbamate (wt%)***
F2356-137-1	0.00	100.00	0.00
F2356-137-2	5.00	95.00	0.00
F2356-137-3	10.20	89.80	0.00
F2356-137-8	15.04	84.96	0.00
F2356-137-4	0.00	80.63	19.37
F2356-137-5	5.00	76.60	18.41
F2356-137-6	10.01	72.55	17.43
F2356-137-7	15.13	68.43	16.44
F2356-142-1	0.00	57.73	42.28
F2356-142-2	5.05	54.81	40.14
F2356-142-3	10.00	51.95	38.05
F2356-142-4	15.03	49.05	35.92
F2356-143-1	0.00	43.35	56.65
F2356-143-2	5.05	41.16	53.78
F2356-143-3	10.07	38.99	50.94
F2356-143-4	14.93	36.88	48.19
F2356-154-1	10.01	25.24	64.75
F2356-154-2	15.13	23.81	61.07

The equation that Aspen uses for mixture viscosity[5] is in the form shown in Equation 2.

Equation 2. Mixture viscosity expression used in Aspen Plus.

$$ln\mu^{l} = \sum X_i ln\mu_i^{*,l} + \sum k_{ij} X_i X_j ln\mu_{ij} + \sum X_i \left[\sum_{j \neq i} X_j (l_{ij} ln\mu_{ij})^{1/3} \right]^3$$

where X_i = mole fraction or weight fraction of component i
k_{ij} = symmetric binary parameter ($k_{ij} = k_{ji}$)
l_{ij} = antisymmetric binary parameter ($l_{ij} = -l_{ji}$)

The binary parameters k_{ij} and l_{ij} allow accurate representation of complex liquid mixture viscosity temperature dependence. Both binary parameters default to zero. Both binary parameters, k_{ij} and l_{ij}, must be specified for each component pair[1]. The first interaction parameter for the Andrade liquid viscosity model is fit to the following expression: $k_{ij} = A + B/T$[1]. The second interaction parameter for the Andrade liquid viscosity model is fit to the following expression: $m_{ij} = C + D/T$[1]. By default, these values are zeros, and they need to be specified. These parameters were regressed using MATLAB to match experimental data. The mixture viscosity equation shown in Equation 2 was modified for 3-component GAP-0/GAP-0 carbamate/water mixtures. Antisymmetric parameters (l_{ij}) were not used, and second binary interaction parameters were added. The final equation for which parameters were regressed is shown in Equation 3.

Equation 3. Mixture viscosity correlation for GAP-0/carbamate/water mixtures to be used in Aspen Plus.

$$\begin{aligned} \ln(\mu_{mix}) = {} & w_{GAP} * \ln(\mu_{GAP}) + w_{H2O} * \ln(\mu_{H20}) + w_{CARB} * \left(a + \frac{b}{T}\right) \\ & + w_{GAP} * w_{H2O} * \left(c + \frac{d}{T}\right) + w_{GAP} * w_{CARB} * \left(e + \frac{f}{T}\right) \\ & + w_{H2O} * w_{CARB} * \left(g + \frac{h}{T}\right) + {w_{GAP}}^2 * {w_{H2O}}^2 * \left(j + \frac{k}{T}\right) \\ & + {w_{GAP}}^2 * {w_{CARB}}^2 * \left(l + \frac{m}{T}\right) + {w_{H2O}}^2 * {w_{CARB}}^2 * (o + \frac{p}{T}) \end{aligned}$$

To understand which coefficients in Equation 3 were important for the regression, a stepwise approach was used to find the best reduced model to fit the data. The result of this stepwise regression approach was a reduced model having 7 coefficients, which are presented in Table 5.

Table 5. Summary of regressed coefficients for viscosity correlation shown in Equation 3.

	Notation in equation	Notation in MATLAB	**Reduced model - FINAL MODEL**			
			Parameter #	**7**		
			Name	Value	p-value	AIC
GAPC/1	a	b1	b1	-6.72842	1.37E-22	-530.146
GAPC/2	b	b2	b2	5228.663	9.24E-62	
GAPC/3	c	b3				
GAP/H2O 1	d	b4	b3	18.09482	7.26E-28	
GAP/H2O 2	e	b5				
GAP/GAPC 1	f	b6	b4	-8.58493	1.17E-24	
GAP/GAPC 2	g	b7				
GAPC/H2O 1	h	b8	b5	-17.6739	1.06E-33	
GAPC/H2O 2	i	b9				
GAP/GAPC 3	j	b10	b6	27.16638	9.42E-16	
GAP/GAPC 4	k	b11				
GAP/H2O 3	l	b12	b7	-57.9421	1.78E-06	
GAP/H2O 4	m	b13				
GAPC/H2O 3	n	b14				
GAPC/H2O 4	o	b15				

Model predictions were validated with experimental data, and results fell within 20% error as shown in Figure 8 for the full range of viscosity values and in Figure 9 for the low end of the viscosity range. The coefficients shown in Table 5 were input into Aspen Plus for use in process model development, analysis, and scale up.

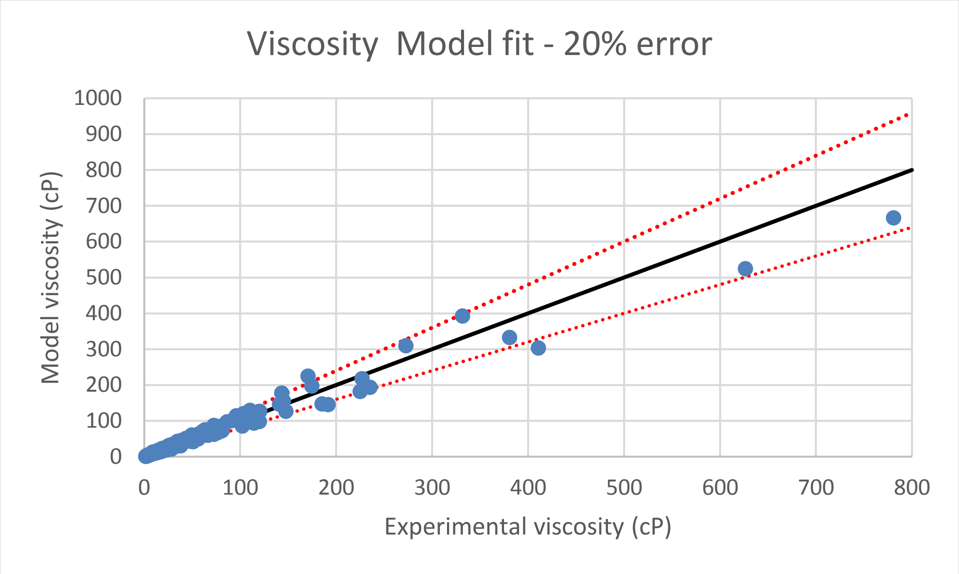


Figure 8. Comparison of Experimental viscosity data with regressed values.

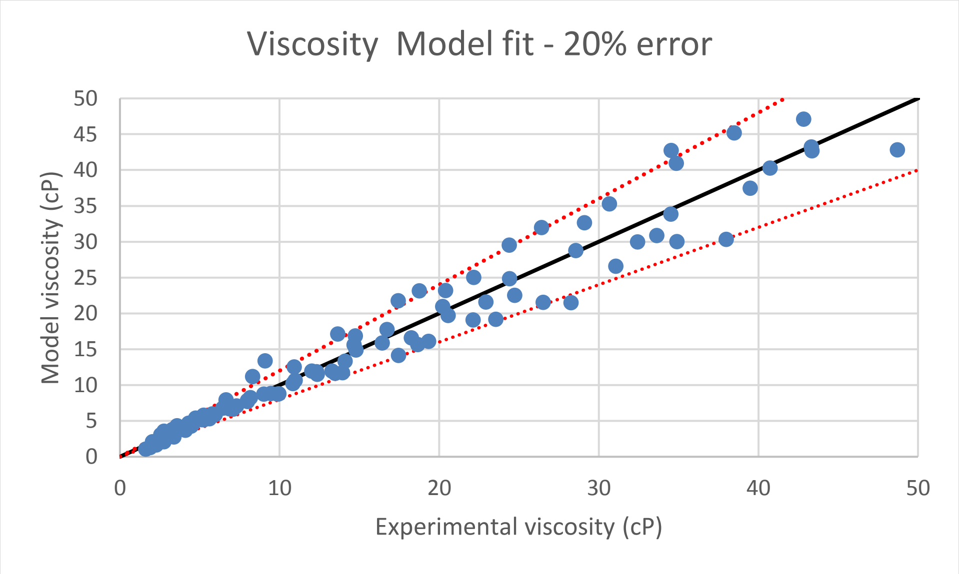


Figure 9. Comparison of experimental viscosity data with regressed values for low viscosity values.

In this analysis, this pump is modeled and costed as a centrifugal pump. Using the improved Aspen Plus model, the flow rate, temperature, and viscosity through the slurry pump were calculated. Additionally, a required pump head was calculated using approximated pipe sizes and lengths in addition to the pressure drops through the required equipment. This data was given to a pump supplier with a request to design and price an appropriate slurry pump. The pump supplier responded with a pump price and required motor size. This information was incorporated into the cost estimates for the new cases. It should be noted that the Aspen Plus model still does not accurately calculate the pump power required for the slurry pump. This can be at least partially attributed to the fact that the model has no information regarding the physical layout or required discharge pressure.

Additional improvements to the Aspen Plus model include incorporating the molar volume in the physical properties as well as correcting the regression of the heat capacity of GAP-0 to better match experimental data.

Desorber Design

The phase-changing aminosilicone solvent has significantly lower vapor pressure, which facilitates conducting desorption in a continuous stirred tank reactor (CSTR) rather than a distillation column. Among the advantages of a CSTR are easier operation and maintenance. In previous work, the CSTR was operated at elevated pressure to reduce CO_2 compression cost. The GAP-0 desorption isotherms (Figure 10) show that high temperatures (>140°C) are needed to desorb CO_2 at elevated pressure. However, at elevated temperatures aminosilicone carbamate readily forms urea byproducts. As a result, the phase-changing CO_2 capture process in this analysis utilizes a single desorber operating at atmospheric pressure. For further description of the CSTR desorber design as applied to the high-pressure concept, see the Preliminary Process and Cost Modeling report[3]. The impact of thermal degradation on CO_2 capture cost is discussed in a later section of this report.

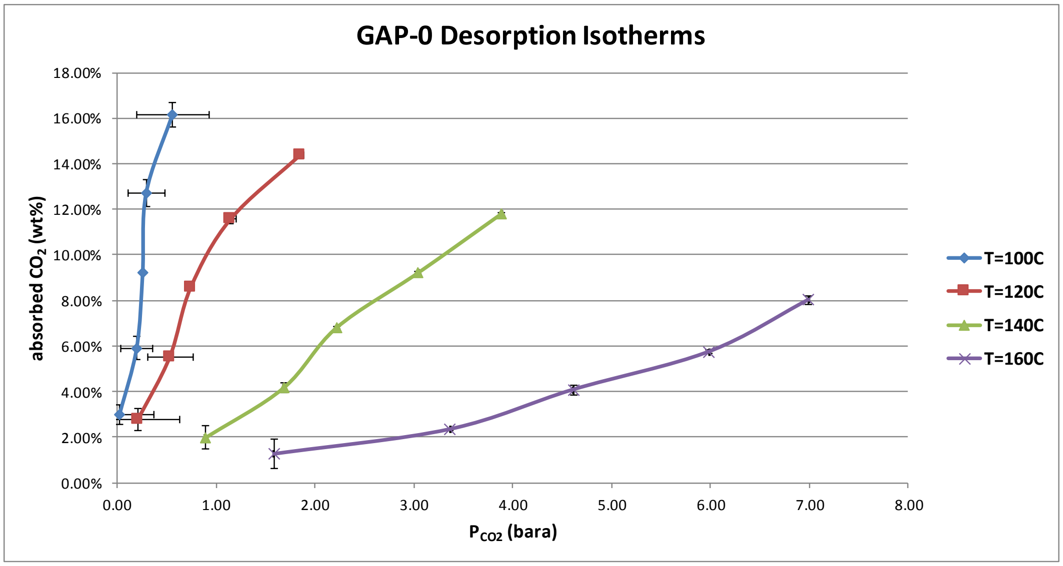


Figure 10. Reaction isotherms for CO_2/phase-changing aminosilicone solvent.

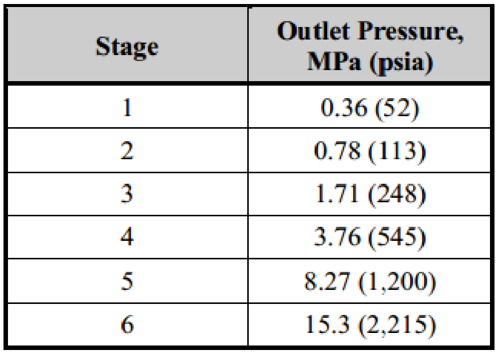
Rich/Lean Heat Exchanger

To recover as much heat as possible from the hot lean solvent leaving the desorber system, this stream will be used to preheat the rich solvent leaving the absorber using a rich/lean heat exchanger. In the process model, the rich/lean heat exchanger is modeled as a shell and tube unit with a constant value of the overall heat transfer coefficient of 75 $Btu/hr \cdot ft^2 \cdot F$. For further description of the rich/lean heat exchanger design process, see the Preliminary Process and Cost Modeling report[3] .

CO_2 Compression

The purpose of the compression train is to deliver a high-purity CO_2 stream at 2215 psia for transportation and storage. The discharge pressures at each compression stage are presented in Table 6.

Table 6. The outlet pressures at each stage in the compression train.



Stage	Outlet Pressure, MPa (psia)
1	0.36 (52)
2	0.78 (113)
3	1.71 (248)
4	3.76 (545)
5	8.27 (1,200)
6	15.3 (2,215)

The atmospheric pressure CO_2 product stream from the desorber is cooled to 90°C to condense and remove any solvent in the stream. It is further cooled to 40°C to condense and remove water from the stream. The resulting CO_2 product stream is compressed and cooled to 48.9°C after each compression stage with cooling water. Any liquid condensate is removed in a vapor/liquid flash separator. Cooling water is supplied from the power plant cooling tower system. The compressors at each stage have a polytropic efficiency of 86% and mechanical efficiency of 98%.

The final CO_2 stream must satisfy the conceptual design limits for enhanced oil recovery as listed in Exhibit 2-1 of the NETL QGESS titled "CO_2 Impurity Design Parameters".[6] Table 7 shows the required specifications for the product CO_2 stream in comparison to the composition of the CO_2 stream for the aminosilicone base-case model. It is clear from Table 7 that the high-pressure CO_2 stream generated by the phase-changing aminosilicone process (Cases 6Q and 6R as defined in Table 1) meets the required specification for CO_2 EOR.

Table 7. Cases 6Q-6R CO_2 stream outlet composition compared to EOR specifications[6].

Component	Unit	Enhanced Oil Recovery specification		CASE 6Q & 6R
		Conceptual design	Range in Literature	
CO_2	Vol %	95	90-99.8	99.8
H_2O	ppm_v	500	20-650	326
N_2	Vol%	1	0.01-2	0.2
O_2	Vol%	0.001	0.001-1.3	0.008
Ar	Vol%	1	0.01-1	0

Basis of Economic Analysis

Capital cost estimations for the carbon capture unit were completed for MEA and the phase-changing aminosilicone cases in order to calculate the first-year COE and the first-year removal cost of CO_2. The annual costs were estimated according to the same basis as described in the Preliminary Process and Cost Modeling report[3], which is presented here for completeness.

Annual cost includes the following items:

- Power plant – capital cost, operating cost, and fuel - The estimated values were compared against DOE Case 11[2]. Further estimates were conducted for a power plant that would be required for 550 MW net power with carbon capture using phase-changing aminosilicone solvent.
- Capital recovery and other fixed charges- The recovery charges are dependent on the Capital Charge Factor (CCF). The CCF used in this study was chosen based on NETL's cost estimation methodology using the case for High risk IOU for five years[6].
- Cost of cooling water- The cost of cooling water from the BBS was used for the non-scaled cases[2]. For the scaled-up cases, the increased cooling water demand increased cooling tower capital cost and operating cost.
- CO_2 transport, storage and monitoring- \$10/tonne as provided by DOE in the cooperative agreement.
- Solvent cost- The solvent cost is consistent with the solvent cost basis used in the Preliminary Process and Cost Modeling report[3].
- Fixed O&M costs- Estimated using a plant on-stream factor of 310.25 days and a charge of \$875/day.
- Maintenance and material cost- Estimated using 1.6% of the material cost.

The details of the calculations are provided below.

Power Plant – Capital Cost, Operating Cost, and Fuel

This cost is the same for all non-scaled cases. It can also be calculated using the following expression:

$$Power\ island\ cost = COE \cdot power\ generated$$

The value of COE used in this expression is equal to 80.95 mils/kWh, from DOE Case 11[2] COE w/o TS&M.

For the scaled-up cases, the cost was estimated using Thermoflow calculations.

Capital Recovery and Other Fixed Charges

The capital recovery was calculated based on the following formula:

$$Capital\ recovery = Capital\ charge\ factor * installed\ CAPEX$$

The capital charge factor (CCF) value is selected based on several factors:

- Type of power plant financial structure (IOU vs. IPP)
- High risk or low risk finance structure
- Capital expenditure period: three years vs. five years.

Table 8 reports capital charge factors for a variety of financial structures[6].

Table 8. Capital charge factors for various financial structures.

Finance Structure	High Risk IOU		Low Risk IOU	
Capital Expenditure Period	Three Years	Five Years	Three Years	Five Years
Capital Charge Factor (CCF)	0.111	0.124	0.105	0.116
Finance Structure	High Risk IPP		Low Risk IPP	
Capital Expenditure Period	Three Years	Five Years	Three Years	Five Years
Capital Charge Factor (CCF)	0.177	0.214	0.149	0.176

The value selected for the phase-changing CO_2 capture process is 12.4%, which corresponds to a high-risk IOU structure with a five-year capital expenditure period.

First year COE was calculated based on the following formula:

$$COE = \frac{total\ annual\ costs}{power\ generated}$$

First year removal cost for CO_2 was calculated using the following expression:

$$Removal\ cost\ \left(\frac{\$}{ton}\right) = \frac{COE_{with\ capture} - COE_{without\ capture}}{lb\ of\ CO2\ separted * power\ generated}$$

Total Cost of Cooling Water

The total cost of cooling water was determined based on the amount of cooling water required as predicted by the Aspen Plus model for the carbon capture process and the cost of cooling water.

CO_2 Transport, Storage, and Monitoring

This cost was calculated based on the amount of CO_2 separated and the cost of transportation, storage, and monitoring (TS&M).

Maintenance Material Costs

The maintenance material costs were calculated from the formula below:

$$\begin{aligned}&\mathit{Maintenance\ material\ costs}\\&\qquad= \mathit{Equipment\ and\ material\ costs} * \mathit{Maintenance\ and\ material\ cost}\ \%\end{aligned}$$

The first-year removal cost of CO_2 was estimated for a supercritical power plant with carbon capture using MEA as a solvent. The results are shown in Figure 11 as compared to DOE Case 12[2]. The values are in good agreement with each other. This analysis validates the methods employed for process and cost analysis, and the result is a reference for comparison with the phase-changing aminosilicone process.

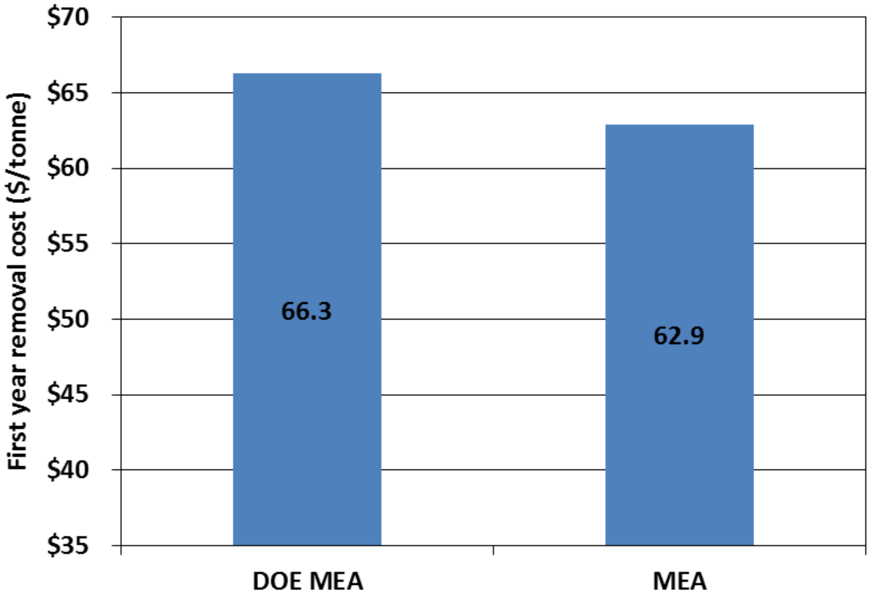


Figure 11. Comparison of first year CO_2 removal cost (2011 basis) using MEA vs. DOE estimated value.

Using this same method, the removal cost was estimated for a supercritical power plant using the phase-changing aminosilicone carbon capture process. Parametric studies were completed assuming constant power plant size per DOE Case 11[2] (550MW before carbon capture). This approach allowed for more efficient exploration of the effect of various process configurations and operating parameters for the CO_2 capture system. Once a best case was identified for the carbon capture unit, the scale was adjusted to achieve 550 MW net power with carbon capture and the CO_2 capture system was heat integrated with the power plant.

For this analysis, the capital cost estimations for the aminosilicone cases were done using Aspen Capital Cost Estimator (ACCE) v8.8 with a cost basis of Q1, 2014. Previous TEAs[3,7] used v7.3.1 (updated in 2011). AspenTech no longer supports v7.3.1 and sometime in early 2017 will no longer support v8.4. During the 3+ years between these versions, AspenTech has made many improvements to their program including changes to the cost basis and default escalation factors. Comparing v7.3.1 to v8.8, there is approximately a 2% increase in the equipment costs and a 3% reduction in the project total cost. This difference was considered to be small, therefore the decision was made to switch to v8.8 moving forward.

The capital costs for Case 6F, the best non-scaled case from the Preliminary Cost Study[7], and Case 6H, the scaled-up case from the Preliminary Cost Study[7], were re-calculated using ACCE v8.8 to allow for direct comparison of the new cases with the previous study. The costs were then adjusted using the CEPCI index to the desired cost basis of 2011.

Economic Parametric Study at 550MW$_{gross}$ Power Plant Scale

The cases evaluated in this study are shown in Table 9. The best case described in the Preliminary Cost Study[7] before scaling to 550MW$_{net}$ was Case 6F, and is the starting point for the cases described herein. Cases 6F and 6J-6P are the parametric study of number of absorption stages, solvent division between absorber stages, and solvent water content. The best of these cases (Case 6J) was scaled to 550MW net power output.

Table 9. CO$_2$ capture cost estimates for 550MW gross power with capture.

Case	# of absorber stages	Lean splits to absorber stages	Desorber T, C	Rich/Lean U, BTU/hr*ft^2*F	SOx conc to CCS, ppmv	Rich solvent	Rich pump	550 MW net w/CO$_2$ Capture?	CO$_2$ removal cost (2011 $ /tonne)	COE w/o TS&M (¢/kWh, 2011 basis)
MEA					10			Y	66.4	13.7
6F*	10	10 equal	140	75	0.25	5wt% of H2O	Slurry Pump	N	60.7	13.1
6J	3	3 equal	140	75	0.25	5wt% of H2O	Slurry Pump	N	59.0	12.9
6K	3	3 equal	140	75	0.25	10wt% of H2O	Slurry Pump	N	64.4	13.7
6L	3	3 equal	140	75	0.25	15wt% of H2O	Slurry Pump	N	73.0	15.3
6M	3	50/25/25	140	75	0.25	5wt% of H2O	Slurry Pump	N	59.1	12.8
6N	3	25/50/25	140	75	0.25	5wt% of H2O	Slurry Pump	N	59.5	12.9
6O	3	25/25/50	140	75	0.25	5wt% of H2O	Slurry Pump	N	60.1	13.0
6P	4	4 equal	140	75	0.25	5wt% of H2O	Slurry Pump	N	64.7	13.6

* Updated from previous TEA to include initial solvent fill cost and use of Aspen Capital Cost Estimator v8.8

The primary focus of the case studies listed in Table 9 relates to the absorber configuration since this was the part of the model that was most changed. In Case 6F from the Preliminary Cost Study[7], the absorber was represented as 10 equal stages as shown in Figure 4. In cases 6J-6L, the absorber was changed from 10 stages to 3 stages and required only 3 stages of intercooling. (See Figure 5.) The effect of changing the amount of water in the rich stream was analyzed. (See Table 10.) The weight percent water was increased in 5% increments from 5 to 15%. The added water decreased the temperature in the absorber so the solvent operated more efficiently. Less solvent was required for the same 90% capture of the CO$_2$ in the flue gas than for the cases with less water. The rich stream temperature was cooler at the higher water levels; therefore, the rich/lean heat exchanger and lean cooler were smaller. However, the added water increased the amount of steam needed in the desorber to reach the same desorber temperature. The size of the CO_2 cooler and the knockout cooler were larger to condense and remove the additional water. The net effect of increasing the solvent water concentration from 5 to 15% was an increase in the cost of CO_2 capture by $13.9/tonne.

Table 10. Absorber Conditions - Cases 6J-6L

	Case 6J - 5 wt% water in Rich Stream			Case 6K - 10 wt% water in Rich Stream			Case 6L - 15 wt% water in Rich Stream		
	Stage 1	Stage 2	Stage 3	Stage 1	Stage 2	Stage 3	Stage 1	Stage 2	Stage 3
Stage split	0.33	0.33	0.34	0.33	0.33	0.34	0.33	0.33	0.34
L/G (Molar Ratio GAP0 to CO2 @ absorber inlet)	0.48	0.73	1.42	0.37	0.56	1.06	0.37	0.55	1.06
Temp of stage °C	70.3	68.4	64.2	62.5	62.0	59.6	59.4	59.2	57.5
water wt% in lean stream to absorber inlet	0.094	0.094	0.094	0.167	0.167	0.167	0.221	0.221	0.221
Temperature of combined rich stream °C	67.7			61.4			58.7		

The next analysis involved changing the solvent splits for the 3 absorber stages. In Case 6J, the lean solvent was split equally between the 3 stages. In Cases M-O, half of the solvent was preferentially sent to one of the stages and the other half was equally split between the remaining two stages. The stage conditions are shown in Table 11.

Table 11. Absorber Conditions – Cases 6J and 6M - 6O

	Case 6J - 5 wt% water in Rich Stream			Case 6M - 5 wt% water in Rich Stream			Case 6N - 5 wt% water in Rich Stream			Case 6O - 5 wt% water in Rich Stream		
	Stage 1	Stage 2	Stage 3	Stage 1	Stage 2	Stage 3	Stage 1	Stage 2	Stage 3	Stage 1	Stage 2	Stage 3
Stage split	0.33	0.33	0.34	1/2	1/4	1/4	1/4	1/2	1/4	1/4	1/4	1/2
L/G (Molar Ratio GAP0 to CO2 @ absorber inlet)	0.48	0.73	1.42	0.72	0.67	1.21	0.38	1.05	1.26	0.40	0.56	1.82
Temp of stage °C	70.3	68.4	64.2	71.8	66.7	63.0	69.6	70.1	63.3	70.2	68.9	66.2
water wt% in lean stream to absorber inlet	0.094	0.094	0.094	0.094	0.094	0.094	0.091	0.091	0.091	0.089	0.089	0.089
Temperature of combined rich stream °C	67.7			68.3			68.3			67.9		

Cases 6J and 6M are very similar. The equipment sizes for both cases are almost identical. There is a slight benefit to case 6J due to the temperature in the absorbers. The first stage in case 6J operates slightly cooler which results in a reduced solvent requirement. In cases 6N and 6O, most of the CO_2 capture takes place in the 2nd and 3rd absorbers respectively. The temperature of the absorber stages stays higher, therefore more solvent flow is required to achieve 90% capture. The net difference between the best case (6J) and the worst of the 4 cases (6O) is $1.1/tonne CO_2.

The previous cases all assumed that 3 stages of absorption were adequate. To substantiate that hypothesis, in Case 6P an extra absorber stage was included and the lean solvent was split equally between the 4 stages. Table 12 shows the molar L/G ratio increasing from stage to stage, indicating that the liquid to the stage is increasing relative to the amount of CO_2. By the 4th stage, the L/G ratio is greater than 5. The stage temperature decreases significantly by the 4th stage also, suggesting a low rate of CO_2 capture in that stage. This indicates that a 4th stage with the associated equipment is not needed.

Table 12. Absorber Conditions - Case 6P

	Case 6P - 4 stage absorber			
	Stage 1	Stage 2	Stage 3	Stage 4
Stage split	1/4	1/4	1/4	1/4
L/G (Molar Ratio GAP0 to CO2 @ absorber inlet)	0.54	0.83	1.58	5.29
Temp of stage °C	73.2	71.0	66.1	47.8
water wt% in lean stream to absorber inlet	0.076	0.076	0.076	0.076
Temperature of combined rich stream °C	65.1			

The costs discussed herein do not account for the effect of thermal degradation on solvent makeup cost. Thermal degradation is addressed later in this document.

Figure 12 and Figure 13 show the plant efficiency and energy penalty due to the CO_2 capture system for the cases shown in Table 9.

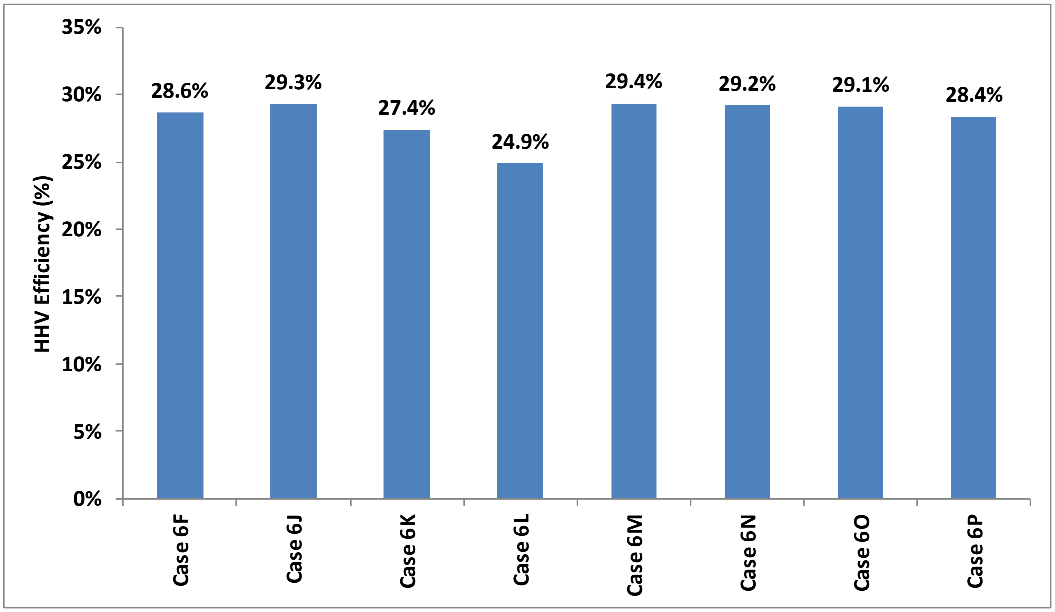


Figure 12. Plant efficiency for each for each of the cases listed in Table 9.

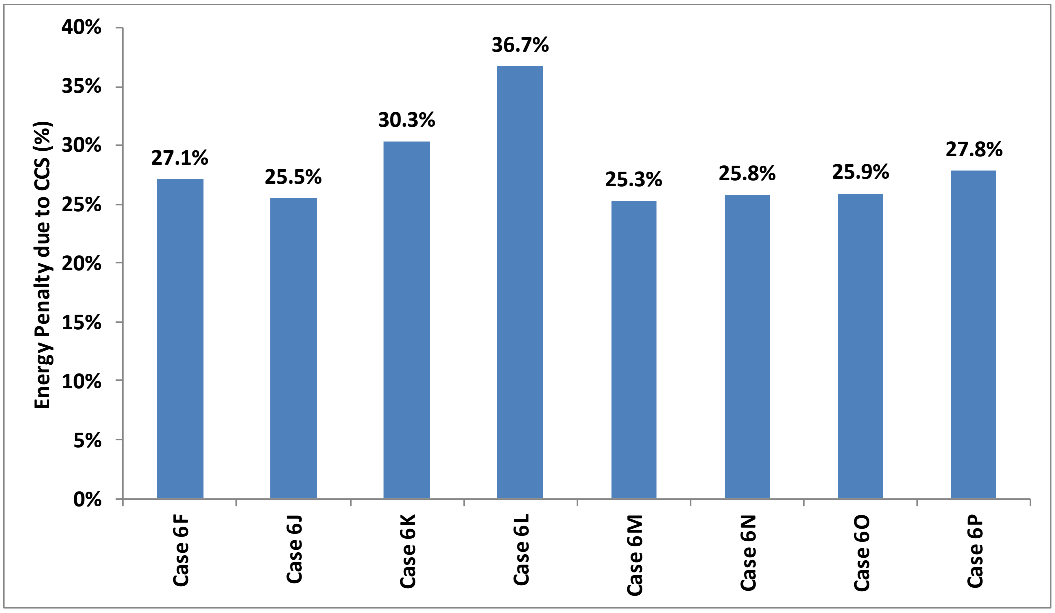


Figure 13. Energy penalty due to CO_2 capture system for each of the cases listed in Table 9.

Economic Analysis at $550MW_{net}$ Power Plant Scale

Because Case 6J from the parametric study had the lowest cost of CO_2 capture, it was scaled up to 550MW net power including the CO_2 capture system to yield Case 6Q. The effect of scaling accounts for the $5.6/tonne decrease in the cost of CO_2 capture compared to the non-scaled case (Case 6J). These cases, compared to MEA, are summarized in Table 13 and Figure 14.

Table 13. CO_2 capture cost estimates for 550MW net power with capture.

Case	# of absorber stages	Lean splits to absorber stages	Desorber T, C	Rich/Lean U, BTU/hr*ft^2*F	SOx conc to CCS, ppmv	Rich solvent	Rich pump	550 MW net w/CO_2 Capture?	CO_2 removal cost (2011 $/tonne)	COE w/o TS&M (¢/kWh, 2011 basis)
MEA					10			Y	66.4	13.7
6H*	10	10 equal	140	75	0.25	5wt% of H2O	Slurry Pump	Y	52.3	12.3
6Q – scaled up	3	3 equal	140	75	0.25	5wt% of H2O	Slurry Pump	Y	53.4	12.3
6R – scaled up, heat integration	3	3 equal	140	75	0.25	5wt% of H2O	Slurry Pump	Y	52.1	12.1
* Updated from previous TEA to include initial solvent fill cost and use of Aspen Capital Cost Estimator v8.8										

As previously mentioned, the best case described in the Preliminary Cost Report[7] was Case 6F. It was scaled up to 550MW net power with CO_2 capture to yield Case 6H. As shown in Figure 14, Case 6Q represents a $1.1/tonne increase in CO_2 capture cost compared to Case 6H. The cost of the equipment and the operating costs for Case 6Q are slightly lower than for case 6H, as shown in Figure 15. The increase in CO_2 capture cost is mainly due to the initial solvent fill, which increased relative to past analyses in part due to a more rigorous basis. Case 6Q uses the newly modified Aspen Plus model while Case 6H uses the previous model. As discussed previously, the new Aspen Plus model is based on actual bench scale data, whereas the previous Aspen Plus model assumed ideal performance. According to the new model based on bench-scale data, approximately 25% more solvent is needed to obtain 90% CO_2 capture.

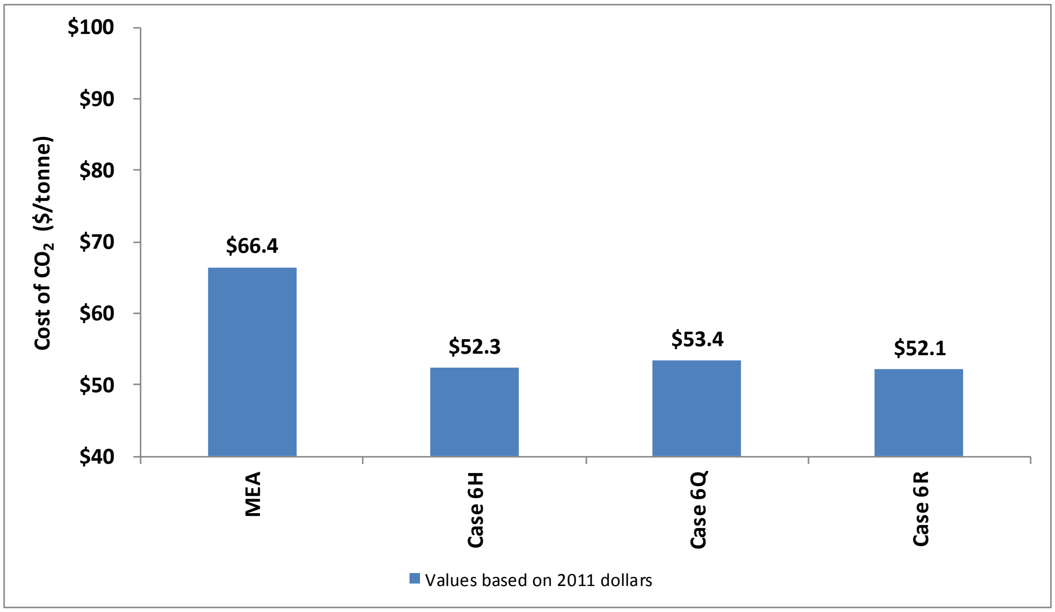


Figure 14. Cost of CO_2 capture for each of the cases listed in Table 13.

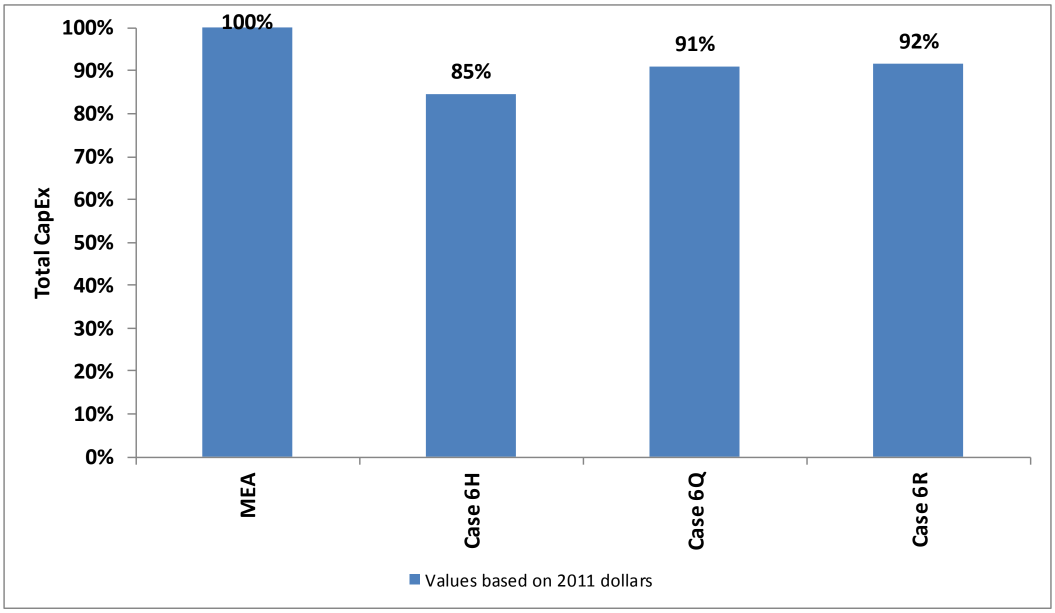


Figure 15. Total capital cost comparison of cases using phase-changing aminosilicone CO_2 capture process vs. DOE Case 12 using MEA, at constant 550 MW_{net} power output.

As shown in Figure 16, the plant efficiency for Case 6Q is 28.3%, compared to 28.4% for DOE Case 12[2], which uses MEA. Case 6R reflects heat integration between the power plant and the phase-changing aminosilicone capture process scaled to 550MW_{net} power output, resulting in overall plant efficiency of 28.3%. While the heat duty associated with desorption is lower for the phase-changing aminosilicone process than for aqueous MEA, this decrease in energy penalty is somewhat offset by an increase in pumping power due to the higher viscosity of the rich slurry. Figure 17 shows the energy penalty due to the CO_2 capture system for these cases. The CO_2 capture system process flow diagram, scaled up to 550 MW net power for Case 6R is presented in Figure 18 and the corresponding stream table is presented in Table 14.

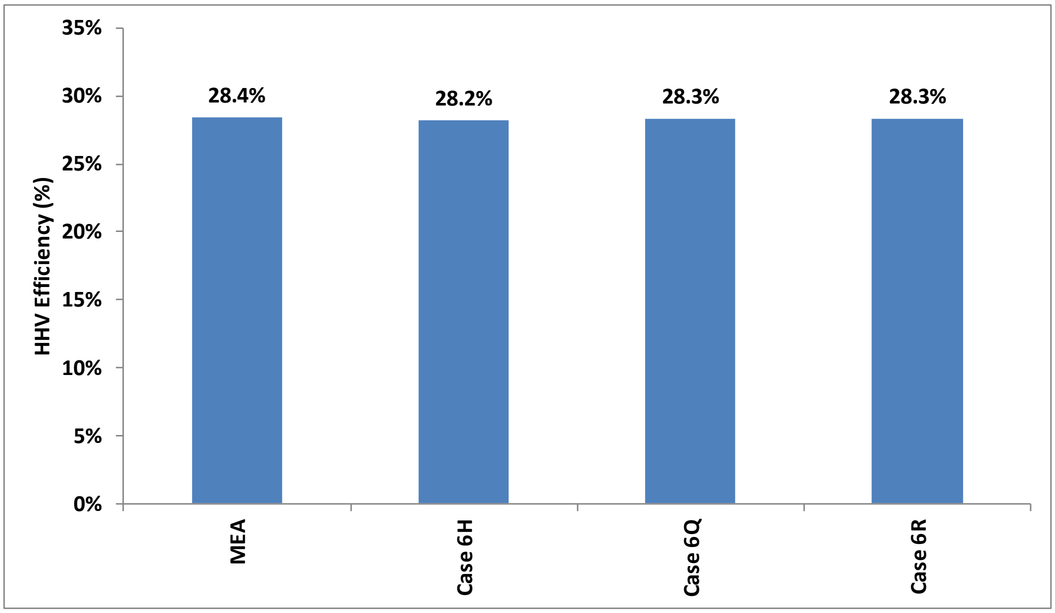


Figure 16. HHV efficiency for each of the cases listed in Table 13.

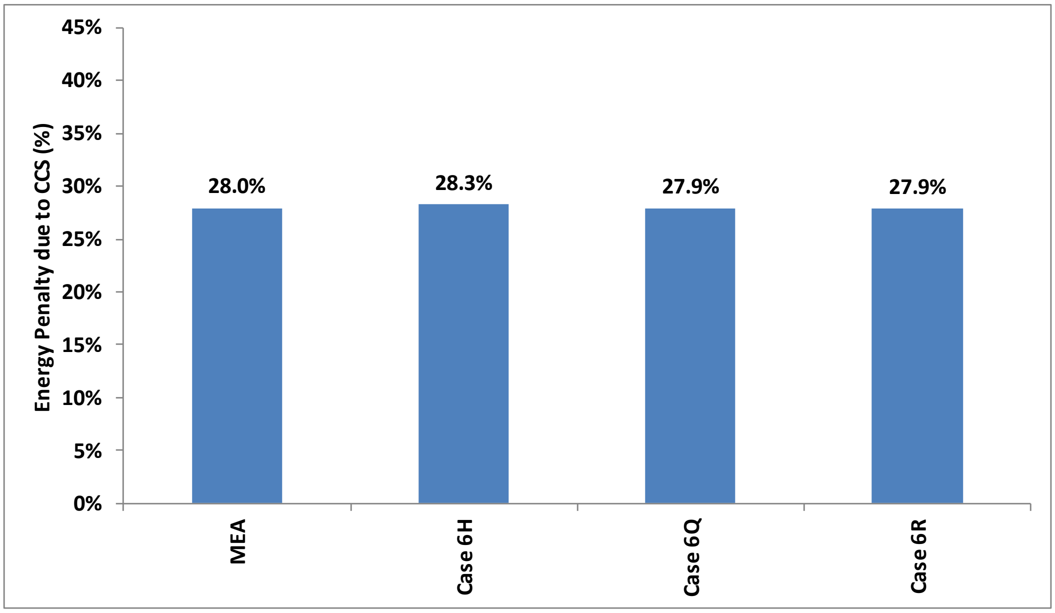


Figure 17. Energy penalty due to CO_2 capture system for each case listed in Table 13.

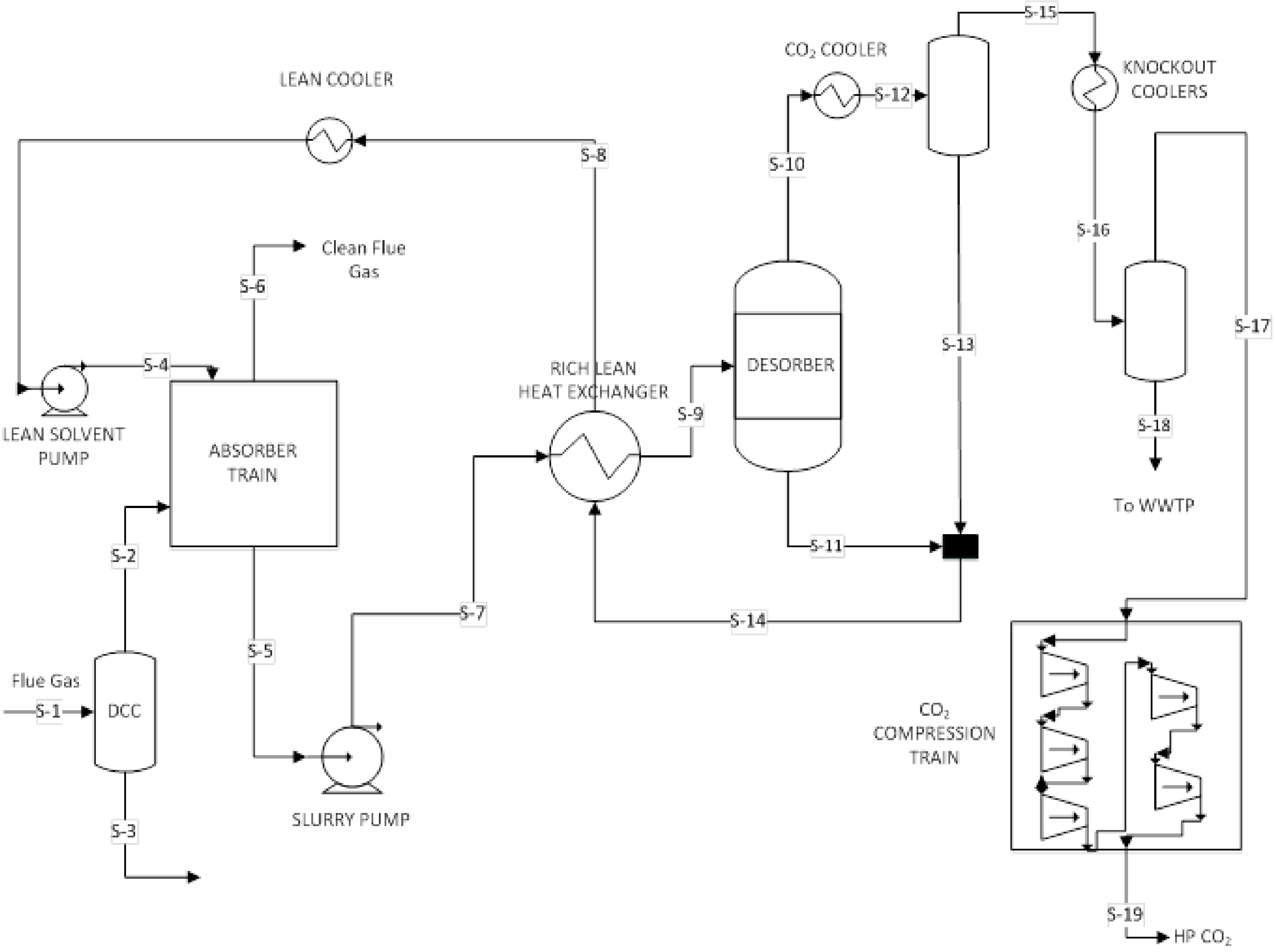


Figure 18. Block flow diagram of CO$_2$ capture system for Case 6R.

Table 14. Stream table for CO_2 capture system for Case 6R.

Stream Number	**S-1**	**S-2**	**S-3**	**S-4**	**S-5**	**S-6**	**S-7**	**S-8**	**S-9**	**S-10**
Mole Fraction										
H_2O	0.152	0.073	0.998	0.593	0.448	0.073	0.448	0.180	0.448	0.515
CO_2	0.135	0.148	0.002	497 PPM	347 PPM	0.017	347 PPM	0.001	347 PPM	0.470
N_2	0.689	0.753	687 PPM	0	601 PPM	0.879	601 PPM	469 PPB	601 PPM	873 PPM
O_2	0.024	0.026	29 PPM	0	25 PPM	0.031	25 PPM	21 PPB	25 PPM	36 PPM
GAP-0	0	0	0	0.382	0.194	143 PPB	0.194	0.769	0.194	0.014
GAP-0CARB	0	0	0	0.024	0.357	trace	0.357	0.049	0.357	93 PPB
Total	1	1	1	1	1	1	1	1	1	1
Total Flow kmol/hr	97,090	88,797	8,292	49,443	36,450	76,026	36,450	24,512	36,450	25,086
Total Flow kg/hr	2,791,990	2,642,180	149,812	5,577,760	5,857,900	2,104,860	5,857,900	5,121,480	5,857,900	840,778
Temperature °C	57.2	40	40	40.6	67.6	40	67.6	78.7	113.7	140
Pressure bar	1.014	1.014	1.014	1.014	1.014	1.014	1.034	1.034	1.034	1.034
Vapor Frac	1	1	0	0	< 0.001	1	0	0	.002	1
Enthalpy kJ/kg	-3,092	-2,530	-15,756	-6765	-6,729	-865	-6,729	-5,922	-6,644	-9,671
Density kg/m^3	1.1	1.2	979.7	869.2	877.9	1.1	878.2	849.1	630.0	1.01
Average MW	28.8	29.8	18.1	112.8	160.7	27.7	160.7	208.9	160.7	33.5

Stream Number	**S-11**	**S-12**	**S-13**	**S-14**	**S-15**	**S-16**	**S-17**	**S-18**	**S-19**
Mole Fraction									
H_2O	0.148	0.515	0.741	0.180	0.502	0.502	0.071	0.988	326 PPM
CO_2	0.001	0.470	0.002	0.001	0.496	0.496	0.928	.011	0.998
N_2	460 PPB	873 PPM	621 PPB	469 PPB	923 PPM	923 PPM	0.002	2 PPM	0.002
O_2	21 PPB	36 PPM	29 PPB	21 PPB	38 PPM	38 PPM	71 PPM	81 PPB	78 PPM
GAP-0	0.799	0.014	0.257	0.769	436 PPM	436 PPM	66 PPB	927 PPM	0
GAP-0CARB	0.052	93 PPB	1.7 PPM	0.049	trace	trace	trace	trace	0
Total	1	1	1	1	1	1	1	1	1
Total Flow kmol/hr	23,163	25,086	1,349	24,512	23,737	23,737	12,574	11,163	11,549
Total Flow kg/hr	5,017,240	840,778	104,237	5,121,480	736,541	736,541	529,937	206,605	507,800
Temperature °C	140	90	90	138.8	90	40	40	40	48.9
Pressure bar	1.034	1.034	1.034	1.034	1.034	1.034	1.034	1.034	152.7
Vapor Frac	0	0.946	0	< 0.001	1	1	1	0	0
Enthalpy kJ/kg	-5,792	-9,812	-7,411	-5,825	-10,152	-10,862	-9,053	-15,502	-8,910
Density kg/m^3	805.3	1.21	825.5	739	1.1	2.3	1.7	979.5	474.9
Average MW	216.6	33.5	77.3	208.9	31.0	31.0	42.2	18.5	44.0

CO_2 Capture Process Utilities

The CO_2 capture process adds additional auxiliary load on coal power plants, and the main contributors are solvent pumps, CO_2 compressors, flue gas blowers, and cooling water fans and pumps. Table 15 shows the power summary for Case 6R of the phase-changing aminosilicone CO_2 capture system. It should be noted that the main feed gas blower is part of the power plant, and only the additional power to increase the flue gas pressure to the required inlet pressure of the CO_2 capture process is shown in Table 15. The cooling tower is also part of the power plant, and its operating and capital costs are included in the power plant costs. Therefore, Table 15 shows only the power for the cooling water pumps, which deliver water from the cooling tower to the CO_2 capture process. CO_2 separation auxiliaries include lean and rich solvent pumps.

Table 15. Utilities summary for Case 6R.

UTILITIES SUMMARY	
AUXILIARY LOAD SUMMARY, kWe	
Feed Gas Blower	919
CO_2 Separation Auxiliaries	3
CO_2 Compression	58,704
Cooling Water Pumps	3,350
TOTAL AUXILIARIES, kWe	62,976
COOLING WATER, tonne/hr	20,227
STEAM, tonne/hr	714

Aqueous Amine CO_2 Capture System Integration with Power Plant

A detailed MEA Aspen Plus model was built to match the analysis in DOE Case 12[2], and was compared to the results of the phase-changing aminosilicone solvent analysis. The main features of the MEA model include an absorber, rich/lean heat exchanger, and a desorber. The baseline MEA case is built from the description given in the BBS[2]. Figure 19 shows a comparison of the plant efficiency reported for DOE Case 12[2] with the plant efficiency calculated using GE Global Research models for MEA as integrated with the power plant described above. The GE Global Research model of the MEA process is in good agreement with DOE Case 12[2], and the plant efficiency of a power plant integrated with a MEA CO_2 capture process integrated with a power plant is 28.4%.

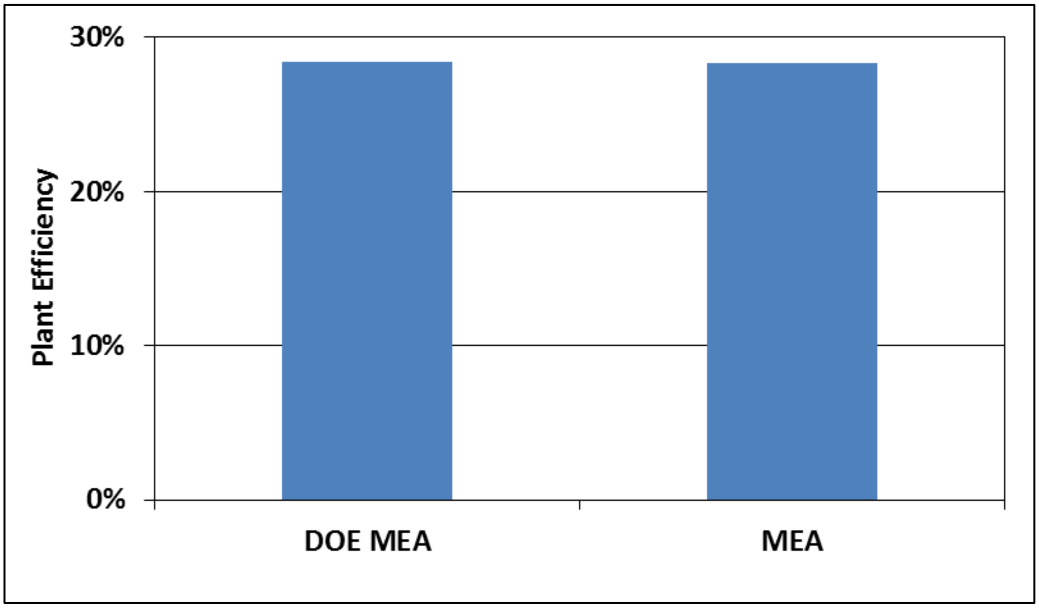


Figure 19. Comparison of estimated plant efficiency of CO_2 capture system using MEA vs. DOE estimated efficiency.

Phase-Changing CO_2 Capture System Integration with Power Plant

In a previous analysis[1] a power plant model was calibrated to DOE Case 11[2] and was used as a starting point to integrate the power plant with the carbon capture process. In the current study, the integrated model resulting from the previous analysis was used as a starting point for the integration of the power plant with the phase changing aminosilicone carbon capture process. Relevant optimization results from the previous study were applied to the phase-changing aminosilicone study, as appropriate.

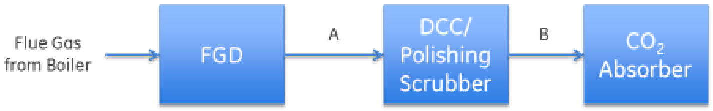
One of the dominant interactions between the power plant and the carbon capture models is the export of process steam for use in the capture plant's desorber. Extracting such a large amount of steam has a significant impact on the design of the power cycle. In the model calibrated to DOE Case 11[2], the low-pressure steam flow was sufficient to require a 4-flow low pressure steam turbine. In the case with carbon capture, a large percentage of the LP steam flow is diverted to the carbon capture plant and thus only a 2-flow low pressure steam turbine is required. The selection of a 2-flow low pressure steam turbine instead of a 4-flow turbine results in a significant difference in steam turbine capital cost (~$60MM). Additionally, the selection of the crossover pressure is heavily influenced by the conditions of steam extraction. The desorber in the carbon capture plant is designed to extract the maximum amount of heat from the process steam by condensing it to a saturated liquid. This sets a minimum steam pressure that can be utilized. If steam is extracted at too low of a pressure, it would not condense at the operating temperatures of the desorber, and a significantly larger extraction of steam would be required. Extracting steam above the minimum pressure doesn't yield significant cost savings, and is worse for the turbine from a performance perspective. As a result, the operating temperature of the desorber directly sets the optimum crossover pressure in the power plant. In this design, crossover pressure was set to 60 psia to align with the desorber operating temperature.

Additionally, feedwater from the power cycle was used as a cooling source in the CO_2 capture process. This arrangement reduces the size of the cooling tower, by limiting how much cooling water is needed in the CO_2 capture process, and benefits the power cycle by pre-warming the feedwater. This allows for multiple low temperature feedwater heaters to be eliminated, therefore less steam needs to be extracted from the steam turbine for feedwater heating, and the overall efficiency of the cycle improves. It was crucial to modify the feed pump turbine condenser so that it operated at the same pressure as the main condenser. This ensured that the mixed feedwater was at as low a temperature as possible, so that it could provide the maximum cooling duty in the CO_2 capture process.

Flue Gas Pretreatment

Sulfur content in the power plant flue gas has a detrimental effect on CO_2 capture equipment and solvents. More stringent flue gas desulfurization requirements in the power plant may be justified based on a reduction in maintenance and material costs for the CO_2 capture plant. A decrease in the permissible sulfur content of flue gas entering the CO_2 capture system increases capital cost and auxiliary load to the power plant.

As in the Preliminary Process and Cost Modeling report[3], solvent losses for all cases were estimated based on an assumed 0.25ppmv SO_2 in the feed flue gas. In this analysis, it is assumed that all SO_2 contained in the flue gas is absorbed by the aminosilicone, which is conservatively consistent with laboratory experiments. In the Bituminous Baseline Study[2], Case 12 utilizes a wet limestone flue gas desulfurization system to reduce the flue gas SO_2 loading to 40ppmv, and a NaOH polishing scrubber having 75-95% efficiency to further reduce sulfur to 10ppmv prior to CO_2 capture. Performance of the FGD in Case 6J was adjusted to deliver 5ppmv SO_2 loading to the polishing scrubber. The polishing scrubber was assumed to have 95% efficiency, which resulted in 0.25ppmv SO_2 entering the CO_2 capture system. This rationale is illustrated in Figure 20. Assuming lower FGD efficiency resulting in a SO_2 concentration of 5ppm, solvent losses due to sulfur deactivation would increase by as much as $12.20/tonne CO_2.



Case	FGD Effluent (Stream A)	DCC Efficiency (75-95% per BBS p. 315)	CO_2 Absorber Inlet (Stream B)
BBS Case 12	40ppm SO_2	75%	10 ppm SO_2
Preliminary Process and Cost Modeling (Case 6F & 6H)	5ppm SO_2	95%	0.25ppm SO_2
Current Analysis (Case 6J-6R)	5ppm SO_2	95%	0.25ppm SO_2

Figure 20. Sulfur removal basis.

Solvent Initial Fill

For this analysis, an attempt was made to more accurately estimate the amount of solvent needed for the initial solvent fill by estimating solvent holdup volumes in the process equipment. The solvent is mainly found in the solvent storage tank, the desorber and the associated piping. The tank sizes were calculated and assumed to be ½ full. Aspen Capital Cost Estimator generated sizes for the system piping. This piping was assumed to be completely full. This total holdup volume was compared to solvent volumes used in installed Advanced Amine process plants.[8] The Advanced Amine plants are different in scale, so the solvent fill was multiplied by the ratio of the power plant gross outputs. The results of the two calculation methods were averaged and compared to the total flow feeding the absorbers in the Aspen Plus model. The average solvent fill is approximately equal to 30 minutes of operation time. The 30 minutes was multiplied by the solvent flow rate and used to calculate the initial solvent fill. This allows a better comparison of the cases and better approximation of the initial fill.

Thermal Degradation and Solvent Makeup

The phase-changing aminosilicone solvent is susceptible to thermal degradation at high temperatures, especially in its CO_2-rich carbamate form. A designed experiment was conducted[9] to quantify this effect, and to identify conditions that limit the degradation mechanism. These experimental results indicate that low temperature and presence of water inhibit the degradation mechanism. However, at the lowest operating temperatures typical of the CSTR, the estimated rate of thermal degradation is estimated to be 1-1.5%/day. This rate of thermal degradation corresponds to as much as \$88/tonne CO_2 for Case 6J.

Results of proof-of-concept experiments for an advanced desorber design suggest that this new desorber can deliver comparable CO_2 desorption rates in the CSTR at temperatures as low as 110°C. The thermal degradation rate at the low temperature and high water content for the advanced desorber is expected to be much lower than that predicted for the CSTR. Preliminary results from the aminosilicone pilot tests[10] indicate that solvent thermal degradation was almost eliminated during the first 170 hours of testing in the advanced desorber, compared to 3% solvent thermal degradation in a comparable period. Further development of this advanced desorber concept is warranted to optimize the phase-changing CO_2 capture process for low cost CO_2 capture.

Power Plant Cost Analysis with CO_2 Capture

Detailed power plant process flow information for each stream in Case 6R with phase-changing aminosilicone CO_2 capture is provided in Table 16. The stream numbers in Table 16 are in reference to the simplified block diagram in Figure 2, and are consistent with the numbering scheme shown for the case without CO_2 capture.

Table 16. Power plant stream table for Case 6R with phase-changing aminosilicone CO_2 capture.

	1	2	3	4	5	6	7	8	9
V-L Mole Fraction									
Ar	0.0093	0.0093	0.0093	0.0093	0.0000	0.0000	0.0088	0.0000	0.0088
CO_2	0.0003	0.0003	0.0003	0.0003	0.0000	0.0000	0.1485	0.0000	0.1485
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0101	0.0101	0.0101	0.0101	0.0000	0.0000	0.0893	0.0000	0.0893
N_2	0.7729	0.7729	0.7729	0.7729	0.0000	0.0000	0.7310	0.0000	0.7310
O_2	0.2074	0.2074	0.2074	0.2074	0.0000	0.0000	0.0202	0.0000	0.0202
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0022	0.0000	0.0022
Total	1.0000	1.0000	1.0000	1.0000	0.0000	0.0000	1.0000	0.0000	1.0000
V-L Flowrate (lbmol/hr)	142,055	142,055	43,642	43,642	-	-	196,676	-	196,676
V-L Flowrate (lb/hr)	4,098,960	4,098,960	1,259,280	1,259,280	-	-	5,852,880	-	5,852,880
Solids Flowrate (lb/hr)	-	-	-	-	549,321	10,901	43,596	43,596	-
Temperature (F)	59	65	59	77	59	59	342	59	342
Pressure (psia)	14.7	15.1	14.7	16.2	14.7	14.7	14.3	14.7	13.8
Enthalpy (BTU/lbm)	-4.3	-3.0	-4.3	0.1	-	-	69.2	-	69.2
Density (lb/ft^3)	0.076	0.078	0.076	0.081	-	-	0.047	-	0.046
V-L Molecular Weight	28.85	28.85	28.85	28.85	-	-	29.76	-	29.76

	10	11	12	13	14	15	16	17	18
V-L Mole Fraction									
Ar	0.0088	0.0000	0.0093	0.0000	0.0000	0.0081	0.0000	0.0000	0.0000
CO_2	0.1485	0.0000	0.0003	0.0000	0.0004	0.1372	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	0.0893	1.0000	0.0101	1.0000	0.9996	0.1577	1.0000	1.0000	1.0000
N_2	0.7310	0.0000	0.7729	0.0000	0.0000	0.6766	0.0000	0.0000	0.0000
O_2	0.0202	0.0000	0.2074	0.0000	0.0000	0.0204	0.0000	0.0000	0.0000
SO_2	0.0022	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
V-L Flowrate (lbmol/hr)	194,494	15,493	3,079	11,440	421	213,657	269,181	224,334	224,334
V-L Flowrate (lb/hr)	5,788,307	279,180	88,848	206,150	7,596	6,155,399	4,850,640	4,042,494	4,042,494
Solids Flowrate (lb/hr)	-	-	-	45,598	77,220	-	-	-	-
Temperature (F)	362.8	59	59	59	132	132	1100	664	1100
Pressure (psia)	15.1	14.7	14.7	14.7	14.7	14.7	3514.7	693.7	655.8
Enthalpy (BTU/lbm)	74.7	27.1	-4.3	-	-	14.9	1495.0	1323.6	1570.5
Density (lb/ft^3)	0.048	62.379	0.076	-	-	0.063	4.319	1.142	0.722
V-L Molecular Weight	29.76	18.02	28.85	-	18.03	28.81	18.02	18.02	18.02

	19	20	21	22	23	24	25
V-L Mole Fraction							
Ar	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
H_2O	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
N_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
O_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
SO_2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
V-L Flowrate (lbmol/hr)	194,864	87,311	19,917	87,635	108,613	108,613	108,613
V-L Flowrate (lb/hr)	3,511,442	1,573,344	358,906	1,579,191	1,957,211	1,573,344	4,850,640
Solids Flowrate (lb/hr)	-	-	-				
Temperature (F)	531	528	531	101	101	322	557
Pressure (psia)	60.0	54.0	60.0	1.0	166.1	133.6	4185.2
Enthalpy (BTU/lbm)	1298.6	1297.3	1298.6	1023.7	69.8	292.6	552.9
Density (lb/ft^3)	0.103	0.093	0.103	0.003	62.010	55.608	47.687
V-L Molecular Weight	18.02	18.02	18.02	18.02	18.02	18.02	18.02

Table 17 summarizes power output from the power plant along with materials consumed during normal operation. It includes a detailed summary of auxiliary loads and how they roll up with the steam turbine power and CO_2 capture and compression loads to impact the total plant net power output and efficiency.

Table 17. Power summary for Case 6R with phase-changing aminosilicone CO_2 capture.

POWER SUMMARY (Gross Power at Generator Terminals, kWe)	
Steam Turbine Power	659,624
Total (Steam Turbine) Power, kWe	**659,624**
Auxiliary Load Summary, kWe	
Boiler Fuel Delivery	4,259
Ash handling	701
Primary Air Fans	1,799
Forced Draft Fans	2,017
Induced Draft Fans	9,833
Baghouse (ESP)	92
Wet FGD	11,970
Econamine FG Plus Auxiliaries	
CO_2 Compression	
Miscellaneous BOP	159
ST Auxiliaries	454
Condensate Pumps	419
Circulating Water Pumps	5,915
Cooling Tower Fans	4,774
Transformer Losses	2,068
BFP Booster Pump	658
Total Auxiliaries, kWe	**45,117**
Net Power, kWe	**614,507**
Net Plant Efficiency (HHV)	33.1%
Net Plant Heat Rate, (BTU/kWh)	10,313
Condenser Cooling duty, (10^6 BTU/hr)	3,215
Consumables	
As-Received Coal Feed, (lb/hr)	543,248
Limestone Sorbent Feed, (lb/hr)	45,612
Thermal Input (kWt)	1,857,348
Raw Water Consumption (gpm)	5,932

The cost summary for the model with CO_2 capture is shown in Table 18. Table 18 shows the calculated annual costs for the power block configured for CO_2 capture. The fixed operating costs and the maintenance and material costs in this case were assumed to be equal to the values in DOE Case 12[2].

Table 18. Equipment cost summary for Case 6R with phase-changing aminosilicone CO_2 capture.

	$ (2011 basis)	$/kW (2011 basis)
Specialized Equipment	$612,900,423	$997
Boiler	$232,510,900	$378
Furnace	$98,147,585	$160
Convective Elements	$65,995,897	$107
Additional Waterwall	$6,430,586	$10
Soot Blowers	$6,286,330	$10
Desuperheaters and Controls	$10,500,845	$17
Air and Flue Gas Ducts	$7,654,276	$12
Coal Pulverisers and Feeders	$25,197,552	$41
FD Fan, PA Fan, ID Fan	$3,979,477	$6.5
Structural Steel, Ladders, Walkways	$3,861,336	$6.3
Rotary Air Heaters	$4,457,014	$7.3
Steam Turbine	$86,960,141	$142
Feedwater Heaters	$9,456,041	$15
Feedwater Heater 1-DA	$910,294	$1.5
Feedwater Heater 2 (2A,2B)	$2,448,927	$4.0
Feedwater Heater 3 (3A, 3B)	$2,830,755	$4.6
Feedwater Heater 4 (4A, 4B)	$3,266,065	$5.3
Water Cooled Condensers	$4,097,311	$6.7
Main Condenser	$ 2,948,860	$4.8
Feed Pump Turbine Condenser	$1,148,451	$1.9
Particulate and Mercury Control	$24,049,049	$39
Flue Gas Desulfurization	$154,562,383	$252
Nitrogen Oxide Control (SCR)	$ 58,437,518	$95
Stack	$11,468,182	$ 19
Continuous Emissions Monitoring System	$638,320	$1.0
Distributed Control System	$1,778,875	$2.9

Transmission Voltage Equipment	$17,132,669	$28
Transformers	$15,247,748	$25
Circuit Breakers	$1,069,121	$1.7
Miscellaneous Equipment	$815,800	$1.3
Generating Voltage Equipment	$11,809,035	$19
Generator Buswork	$6,168,027	$10
Circuit Breakers	$5,078,674	$8.3
Miscellaneous Equipment	$562,335	$0.9

Other Equipment	$88,796,890	$145
Pumps	$15,055,588	$25
Boiler Feed Pump (+ Turbine)	$9,647,957	$16
Boiler Feed Booster Pump	$177,837	$0.3
Condenser C.W. Pump	$3,377,904	$5.5
Condensate Forwarding Pump	$260,537	$0.4
Condenser Vacuum Pump	$410,268	$0.7
Aux Cooling Water Pump (Closed Loop)	$ 45,069	$0.1
Treated Water Pump	$7,337	$0.01
Diesel Fire Pump	$175,847	$0.3
Jockey Fire Pump	$5,273	$0.01
Demin Water Pump	$14,525	$0.02
Raw Water Pumps	$ 35,717	$0.1
Aux Cooling Water Pump (Open Loop)	$45,069	$ 0.1
Startup Boiler Feed Pump	$852,249	$1.4
Tanks	$ 556,991	$0.9
		$ -
Demin Water	$60,427	$0.1
Raw Water	$308,541	$0.5
Neutralized Water	$41,835	$0.1
Acid Storage	$11,566	$0.0
Caustic Storage	$11,566	$0.0
Dedicated Fire Protection Water Storage	$123,056	$0.2
Cooling Tower	$13,478,927	$22
Auxiliary Cooling Water Heat Exchanger	$158,145	$0.3
Steam Turbine Crane	$1,428,499	$2.3

Station Instrument Air Compressors	$972,645	$1.6
General Plant Instrumentation	$457,280	$0.7
Medium Voltage Equipment	$8,822,512	$14
Transformers	$1,278,283	$2.1
Circuit Breakers	$594,687	$1.0
Switchgear	$1,994,519	$3.2
Motor Control Centers	$4,534,919	$7.4
Miscellaneous	$ 420,105	$0.7
Low Voltage Equipment	$2,307,535	$3.8
Transformers	$732,721	$1.2
Circuit Breakers	$783,583	$1.3
Motor Control Centers	$681,361	$1.1
Miscellaneous	$109,871	$0.2
Coal Handling Equipment	$34,878,181	$57
Ash Handling Equipment	$6,452,162	$10
Miscellaneous Equipment	$ 4,228,423	$7

Civil	$122,681,835	$200
Site Work	$20,155,011	$33
Excavation and Backfill	$7,648,289	$12
Concrete	$93,686,614	$152
Roads Parking and Walkways	$1,191,922	$1.9

Mechanical	$302,206,996	$492
On Site Transportation and Rigging	$11,669,295	$19
Equipment Erection and Assembly	$203,128,412	$331
Piping	$84,916,918	$138
Steel	$2,492,370	$4.1

Electrical Assembly and Wiring	$27,133,213	$44
Controls	$16,303,255	$27
Assembly and Wiring	$ 10,829,958	$18

Buildings and Structures	$22,997,477	$37
Boiler House and Turbine Hall	$20,913,528	$34
Administration Control Room, Machine Shop, Warehouse	$2,056,980	$3.3
Guard House	$26,969	$0.04

Engineering and Plant Startup	$57,728,319	$94
Engineering	$46,721,241	$76
Start Up	$11,007,078	$18

Totals		
Subtotal Contractor's Internal Cost	$1,234,445,153	$2,009
Contractors Soft & Misc Costs	$237,145,173	$386
Subtotal Contractor's Price	$1,471,590,326	$2,395
Owner's Soft and Misc Costs	$288,782,684	$470
Total Owner's Cost	$1,760,373,010	$2,865

Table 19 shows the calculated annual costs for the power block configured for CO_2 capture. The fixed operating costs and the maintenance and material costs in this case were assumed to be equal to the values in DOE Case 12[2].

Table 19. Annual cost summary for Case 6R with phase-changing aminosilicone CO_2 capture.

			Annual Cost $ (2011 basis)	Annual Unit Cost $/kWh-net (2011 basis)
Fixed Operating Costs			$61,577,581	$0.01144
Maintenance Material Costs			$18,298,142	$0.00340
	Consumption / day	Unit Cost		
Water (/1000 gallons)	4,271	1.67	2,213,010	$0.00041
Chemicals				
MU & WT Chem.(lbs)	20,673	0.27	$1,731,754	$0.00032
Limestone (ton)	644	33.48	$6,688,653	$0.00124
Ammonia (19% NH3) ton	98	330	$10,050,143	$0.00187
Subtotal Chemicals			$18,470,550	$0.00343
Other				
SCR Catalyst (m3)	0.41	5776	$736,904	$0.00014
Subtotal Other			$ 736,904	$0.00014
Waste Disposal				
Total Ash (ton)	632	25.11	$4,926,177	$0.00092
Subtotal Waste Disposal			$4,926,177	$0.00092
Total Variable Operating Costs			$26,346,641	$0.00489
Fuel (ton)	6519	68.60	$138,744,428	$0.02577

Table 20 details the energy flows in and out of the power plant integrated with phase-changing aminosilicone CO_2 capture.

Table 20. Energy balance for power plant for Case 6R with phase-changing aminosilicone CO_2 capture.

	HHV	Sensible + Latent Heat	Power	Total
Heat In (MMBTU/hr)				
Coal	6352			6352
Ambient Air		71		71
FGD Water		29		29
FGD Oxidation Air		7		7
Process Return		745		745
Totals	6352	852		7203
Heat Out (MMBTU/hr)				
Bottom Ash		6		6
Fly Ash + FGD Ash		2		2
Flue Gas		810		810
Unburned Carbon		18		18
Boiler Losses		56		56
Fuel Delivery Losses		3		3
Main Condenser		1511		1511
BFPT Condenser		352		352
Process Extraction		2168		2168
Steam Piping Losses		14		14
ST/Generator Mech/Elec/Gear Losses		25		25
BFPT Mech Losses		1		1
Pumps Mech/Elec Losses		3		3
Fans Mech/Elec Losses		5		5
FGD Energy Losses		42		42
Misc Losses and Auxiliaries		89		89
Net Power			2097	2097
Totals	0	5106	2097	7204

Table 21 shows the air emissions for Case 6R.

Table 21. Air emissions for Case 6R with phase-changing aminosilicone CO_2 capture.

	lb/hr (stack)
SO_2	~0
NO_x	~0
Particulates	17.25
Hg	~0
CO_2	127,739

The carbon balance for Case 6R is shown in Table 22.

Table 22. Carbon balance for Case 6R with phase-changing aminosilicone CO_2 capture.

Carbon In, (lb/hr)		Carbon Out (lb/hr)	
Coal	346,321	Stack Gas	34,838
Air (CO_2)	661	FGD Product	699
FGD Reagent	5,474	CO_2 Product	304,893
Total	**352,455**	**Total**	**340,430**

The sulfur balance for Case 6R is shown in Table 23.

Table 23. Sulfur balance for Case 6R with phase-changing aminosilicone CO_2 capture.

Sulfur In, (lb/hr)		Sulfur Out (lb/hr)	
Coal	13,683	FGD Product	13,650
		Stack Gas	0
		Waste Solvent	33
Total	**13,683**	**Total**	**13,683**

Table 24 summarizes the pieces of equipment that contribute to the total water consumption in the power plant with phase-changing aminosilicone CO_2 capture. The major demand for water for this process is for cooling water, which is used for intercooling between stages of CO_2 separation and CO_2 compression. In the baseline design this intercooling is done with water from the cooling tower and all the heat extracted from the separation and compression processes is returned to the cooling tower.

Table 24. Water consumption for Case 6R with phase-changing aminosilicone CO_2 capture.

Water Use	Water Consumption (gpm)
Phase-Changing Aminosilicone CO_2 Capture	2018
FGD Makeup	558
Cooling Tower	5,163
Total	**7,739**

The detailed process flow diagram of the complete power plant integrated with phase changing aminosilicone CO_2 capture is shown in Figure 23.

First year COE was calculated (with and without TS&M) as shown in Figure 21 and Figure 22. Case 6R COE w/o TS&M is 11.8 cents/kWh as compared to 13.73 cents/kWh for the MEA carbon capture system. When TS&M is included in the analysis, Case 6R COE is 12.8 cents/kWh vs. 14.73 cents/kWh for the MEA carbon capture system.

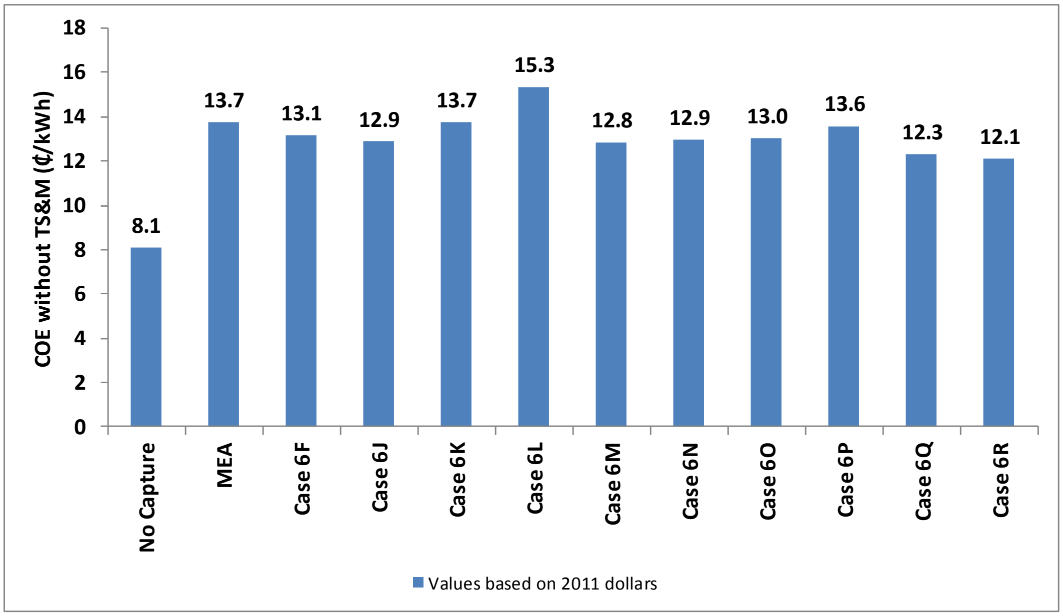


Figure 21. Cost of electricity without TS&M for phase-changing aminosilicone cases.

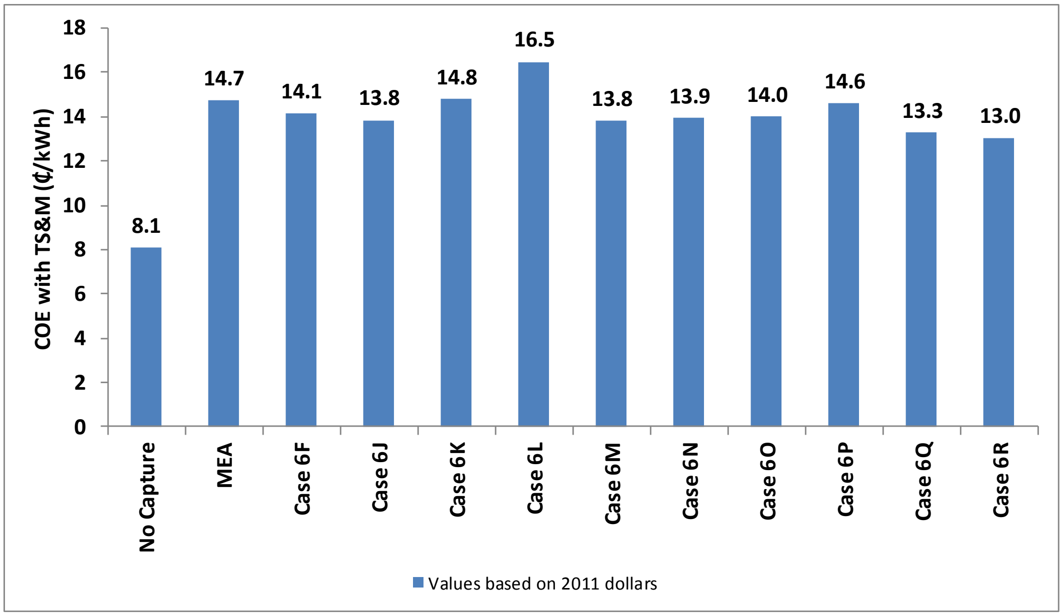
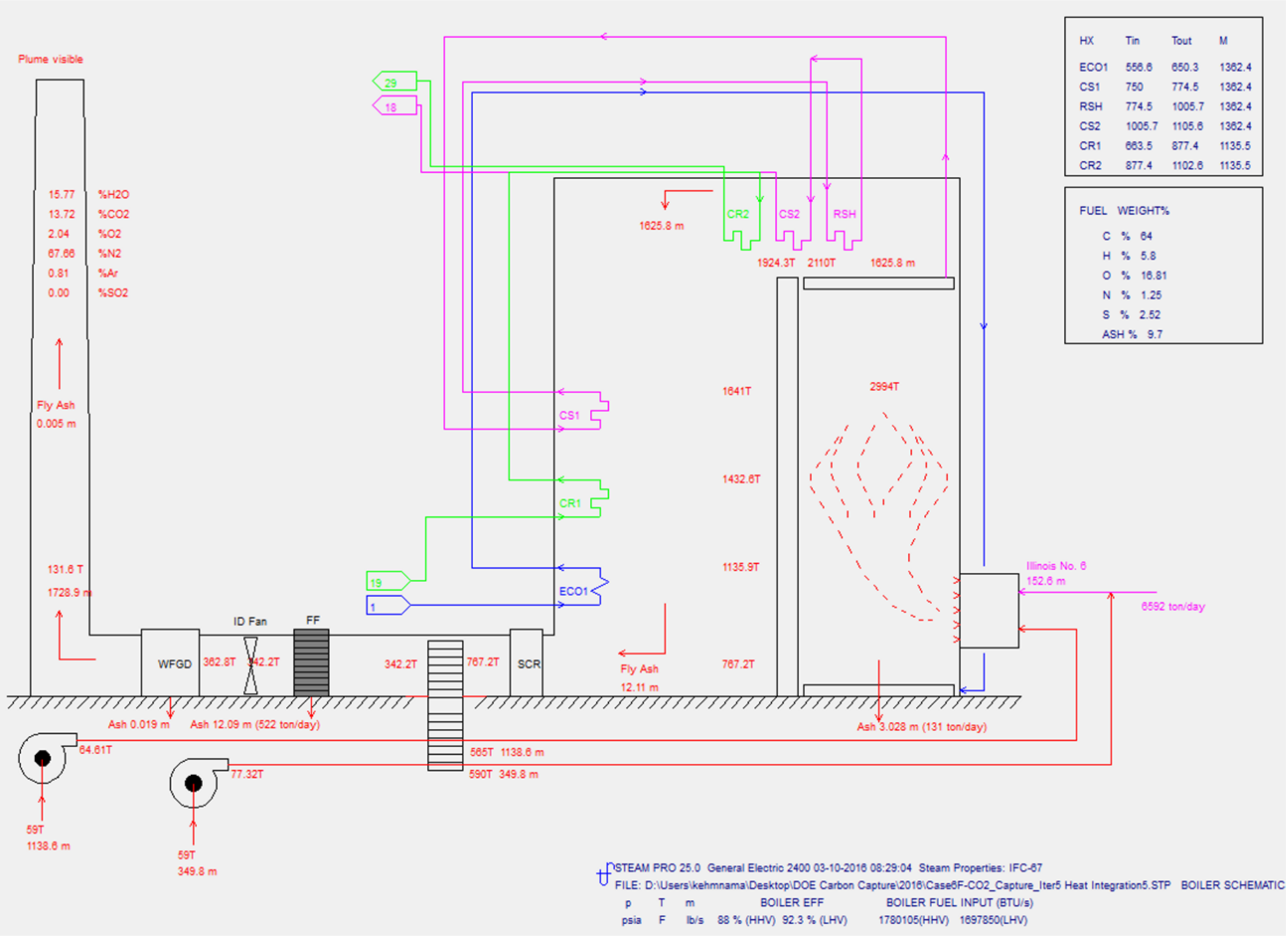


Figure 22. Cost of electricity with TS&M for phase-changing aminosilicone cases.



HX	Tin	Tout	M
ECO1	556.6	650.3	1362.4
CS1	750	774.5	1362.4
RSH	774.5	1005.7	1362.4
CS2	1005.7	1105.6	1362.4
CR1	663.5	877.4	1135.5
CR2	877.4	1102.6	1135.5

FUEL	WEIGHT%	
C	%	64
H	%	5.8
O	%	16.81
N	%	1.25
S	%	2.52
ASH	%	9.7

STEAM PRO 25.0 General Electric 2400 03-10-2016 08:29:04 Steam Properties: IFC-67
FILE: D:\Users\kehmnama\Desktop\DOE Carbon Capture\2016\Case6F-CO2_Capture_Iter5 Heat Integration5.STP BOILER SCHEMATIC

p	T	m	BOILER EFF	BOILER FUEL INPUT (BTU/s)
psia	F	lb/s	88 % (HHV) 92.3 % (LHV)	1780105(HHV) 1697850(LHV)

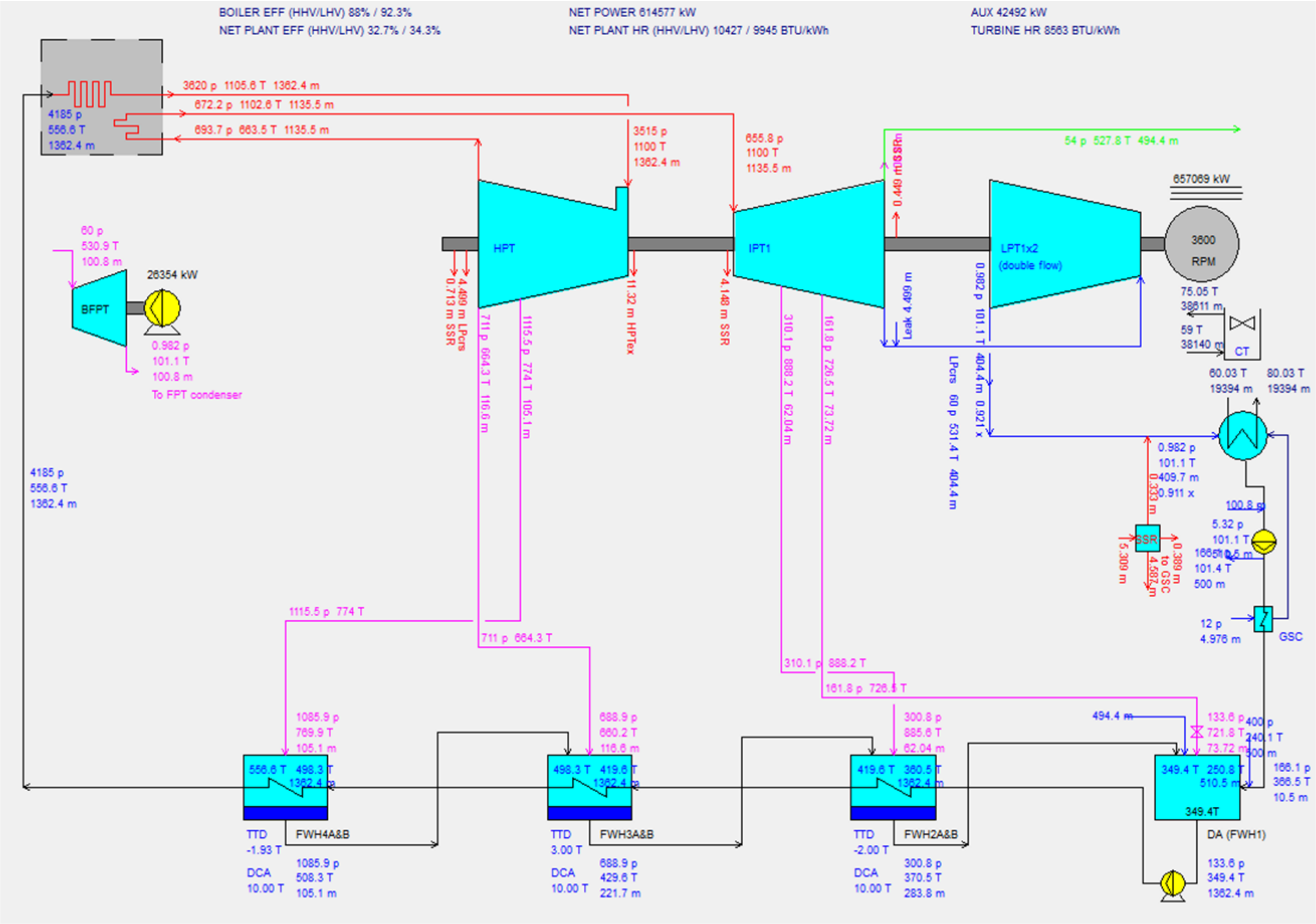


Figure 23. Power plant model for Case 6R with phase-changing aminosilicone CO_2 capture.

Further cost reduction is possible from Case 6R, as shown in Figure 24, due to selection of less costly materials of construction, use of an advanced desorption process, and reducing the material cost of the aminosilicone solvent.

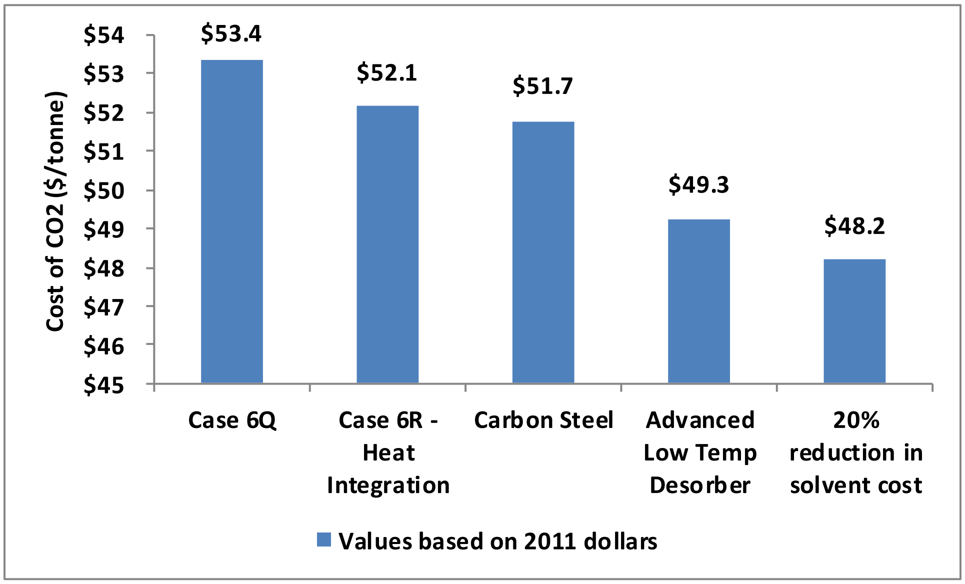


Figure 24. Sensitivity of CO_2 capture cost to materials of construction, working capacity, and solvent material cost.

Capital cost may be reduced through selection of less costly materials of construction for all process units. Experiments to screen the efficacy of additive inhibitors[11] indicated that the carbon steel corrosion rate of GAP-0/carbamate/water mixtures in the presence of inhibitors was lower than that for additive-free aqueous amines. Thus, use of carbon steel equipment may be justified.

Capital and operating cost may be reduced through optimization of the process unit operations for low temperature desorption. For example, implementing a desorption unit operation that enables high desorption rate at low temperature may enable an increase in the working capacity of the solvent compared to that obtained for Case 6R, with less thermal degradation. Further analysis of this concept will be included in an update to the Techno-Economic Analysis following bench scale testing of an advanced low-temperature desorber. The data shown in Figure 24 is currently the best estimate of the impact of these process modifications to CO_2 capture cost using a CSTR desorber.

As shown in Figure 15, the overall capital cost of the phase-changing CO_2 capture process is about 8% lower than that of the MEA process. This overall capital cost includes the initial solvent fill for each process. As shown in Figure 25, a substantial portion of the overall capital cost for the phase-changing CO_2 capture process is due to the initial solvent fill. Ignoring the initial solvent fill, the first-year removal cost of CO_2 for the phase-changing aminosilicone carbon capture process is $47.5/tonne of CO_2 (based on Case 6R), compared to $65.9/tonne of CO_2 when MEA is used. The capital equipment cost (excluding initial solvent fill) of the phase-changing aminosilicone process is about 32% less than that of the benchmark MEA process. Further cost reduction of the aminosilicone process could be achieved by working with silicone suppliers to further reduce the material cost of the solvent and/or by reducing the amount of solvent required for the initial fill, e.g., by increasing the solvent working capacity.

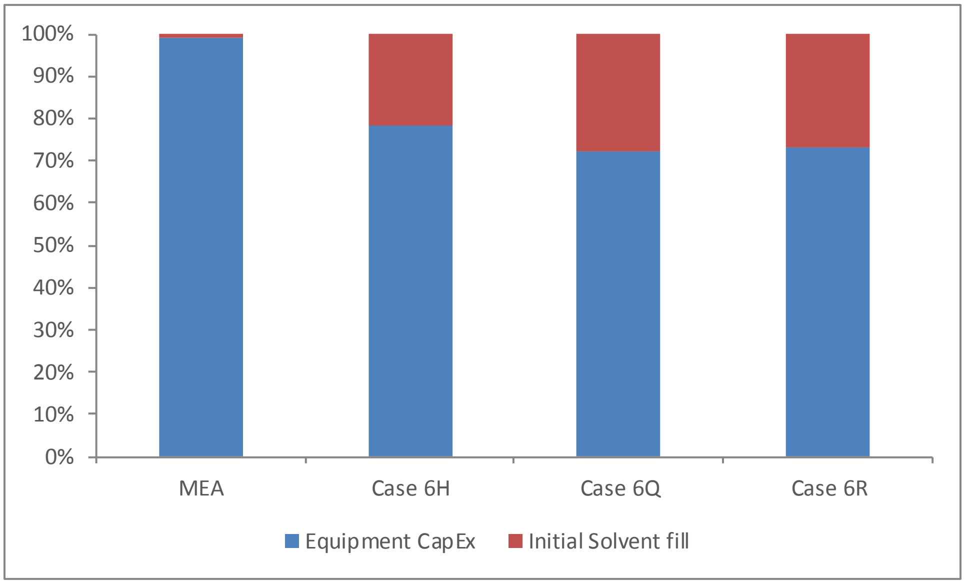


Figure 25. Contribution of initial solvent fill to total capital cost of CO_2 capture systems.

Aminosilicone Process Advantage over MEA

As shown in Table 13, the first-year CO_2 removal cost for the phase-changing CO_2 capture process is $52.1/tonne, compared to $66.4/tonne for the aqueous amine process. The phase-changing CO_2 capture process is less costly than MEA because of advantageous solvent properties that include higher working capacity, lower corrosivity, lower vapor pressure, and lower heat capacity.

- Higher working capacity reduces the solvent flow rate required for 90% capture, resulting in smaller equipment.
- Lower solvent corrosivity allows for selection of less costly materials of construction (i.e., carbon steel) for key unit operations, including the rich/lean heat exchanger.
- Lower vapor pressure results in reduced solvent losses from the system.
- Lower heat capacity increases the energy efficiency of the process by decreasing the sensible heat duty in the desorber and rich lean heat exchanger. As a result, less steam is required for the CO_2 capture system.

These advantageous properties result in a 32% reduction in equipment capital cost, and 8% reduction in overall capital cost for the phase-changing CO_2 capture process compared to MEA. These costs presume that the risk of solvent thermal degradation can be mitigated. Future work is focused on developing technology that addresses this critical risk.

Future Work

In the coming year, a bench-scale advanced, low-temperature desorber will be built and integrated with the bench scale spray absorber. With this new system, experimental data will be generated to quantify the performance of the new desorber. This data will be used to develop an Aspen Plus model of the new process utilizing the absorber model described in this report. The advanced, low-temperature desorber is anticipated to deliver lower thermal degradation rate and higher solvent working capacity compared to the CSTR process. The cost impact of these parameters will be addressed with the advanced desorber in an updated techno-economic analysis.

Conclusions

System and economic analysis for the phase-changing aminosilcone CO_2 capture process integrated with a pulverized coal (PC) boiler demonstrates that the aminosilicone process has significant advantages relative to an MEA-based system. The aminosilicone solvent is 1,3-bis(3-aminopropyl)-1,1,3,3-tetramethyldisiloxane (GAP-0). For comparison purposes, the report also shows results for a carbon-capture unit based on a conventional approach using monoethanolamine (MEA).

The first-year CO_2 removal cost for the phase-changing CO_2 capture process is \$52.1/tonne, compared to \$66.4/tonne for the aqueous amine process. The phase-changing CO_2 capture process is less costly than MEA because of advantageous solvent properties that include higher working capacity, lower corrosivity, lower vapor pressure, and lower heat capacity. The phase-changing aminosilicone process has approximately 32% lower equipment capital cost compared to that of the aqueous amine process. However, this solvent is susceptible to thermal degradation at CSTR desorber operating temperatures, which could add as much as \$88/tonne to the CO_2 capture cost associated with solvent makeup. Future work is focused on mitigating this critical risk by developing an advanced low-temperature desorber that can deliver comparable desorption performance and significantly reduced thermal degradation rate.

List of Acronyms and Abbreviations

ACCE = Aspen Capital Cost Estimator

BBS = Bituminous Baseline Study
CSTR = Continuous Stirred Tank Reactor
DOE = (US) Department of Energy
EOR = Enhanced Oil Recovery
FGD = Flue Gas Desulfurization
GAP-0 = 1,3-bis(3-aminopropyl)-1,1,3,3-tetramethyldisiloxane
GAPC = the CO_2-rich carbamate form of GAP-0
MEA = MonoEthanolAmine
NETL = National Energy Technology Laboratory
PC = Pulverized Coal
SCR = Selective Catalytic Reduction
TS&M = Transportation, Storage, and Monitoring

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