Bench-Scale Silicone Process for Low-Cost CO_2 Capture

Final Technical and Economic Feasibility Report

Topical Report

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DE-FE0007502 Task 8.2

Disclaimer:

Abstract:

This report presents system and economic analysis for a carbon capture unit which uses an amino-silicone solvent for CO_2 capture and sequestration (CCS) in a pulverized coal (PC) boiler. The amino-silicone solvent is based on GAP-1 with tri-ethylene glycol (TEG) as a co-solvent. For comparison purposes, the report also shows results for a CCS unit based on a conventional approach using mono-ethanol amine (MEA).

At a steam temperature of 395 °C (743 °F), the CCS energy penalty for amino-silicone solvent is only 30.4% which compares to a 35.9% energy penalty for MEA. The increase in COE for the amino-silicone solvent relative to the non-capture case is between 98% and 103% (depending on the solvent cost) which compares to an ~109% COE cost increase for MEA. In summary, the amino-silicone solvent has significant advantages over conventional systems using MEA.

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Executive Summary:

This report presents system and economic analysis for a carbon capture unit which uses an amino-silicone solvent for CO_2 capture and sequestration (CCS) in a pulverized coal (PC) boiler. The amino-silicone solvent is based on GAP-1 with tri-ethylene glycol (TEG) as a co-solvent. For comparison purposes, the report also shows results for a CCS unit based on a conventional approach using mono-ethanol amine (MEA).

Aspen Plus^{TM} models were developed for both the MEA and amino-silicone solvent-based CO_2 separation units to calculate the mass and energy balances and system performance. The models account for steam load for the CO_2 separation units and parasitic loads for solvent pumps, CO_2 compressors, and cooling water pumps.

Capital costs were estimated by the Aspen^{TM} Cost Estimator program and a relative cost comparison between the two configurations is presented. The energy penalty for the plant and cost of electricity (COE) were calculated using the assumptions specified by the Department of Energy (DOE) in the contract (DE-FE0007502).

At a steam temperature of 395 °C (743 °F), the CCS energy penalty for amino-silicone solvent is only 30.4% which compares to a 35.9% energy penalty for MEA. At a lower steam temperature of 204 °C (400 °F), the energy penalty for the amino-silicone solvent is 29.0%.

At a steam temperature of 395 °C (743 °F), the increase in COE for amino-silicone solvent relative to the non-capture case is between 98% and 103% (depending on the solvent cost) which compares to an ~109% COE cost increase for MEA. At a lower steam temperature of 204 °C (400 °F), the increase in COE for the amino-silicone solvent is between 95% and 100%. In summary, the amino-silicone solvent has a significant advantage over conventional systems using MEA.

Completion of Task 8.2: Final Technical and Economic Feasibility Study:

Process Description

The pulverized coal (PC) plant and CO_2 separation unit based on mono-ethanol amine (MEA) is described in Case 10 of the DOE report titled "Cost and Performance Baseline for Fossil Energy Plants, DOE/NETL-2007/1281, Volume 1: Bituminous Coal and Natural Gas to Electricity Final Report, May 2007".

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. The flue gas is then sent through the carbon dioxide separation unit before venting the flue gas.

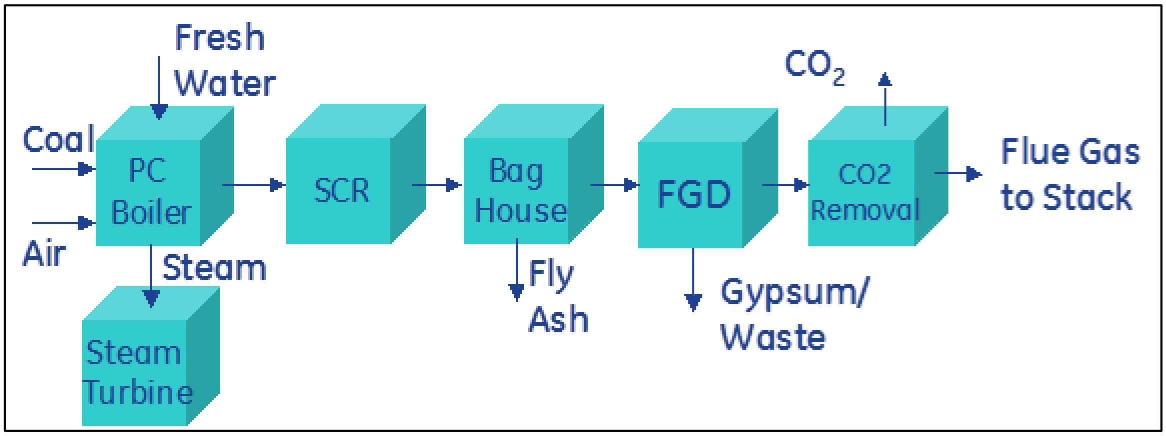


Figure 1. System block diagram.

The MEA and GAP-1/TEG CO_2 separation units utilize four key processes, CO_2 absorption, CO_2 desorption, sorbent 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, as shown in Figure 2. The lean sorbent enters the absorber at 40 °C (104 °F) and captures most of the CO_2 from the flue gas and the rich sorbent leaves the absorber. The CO_2 absorption increases the temperature of the sorbent. The absorber is operated at 66-82 °C (150-180 °F) and at atmospheric pressure.

The rich sorbent from the absorber is fed to the rich-lean heat exchanger and heated up before being fed to the desorber (stripper) for separation of the absorbed CO_2. A 5.6-11.7 °C (10-30 °F) approach is assumed for this rich-lean heat exchanger. This is defined as the hot fluid outlet temperature minus the cold fluid inlet temperature. The lean sorbent from the desorber is passed through the other side of the rich-lean heat exchanger.

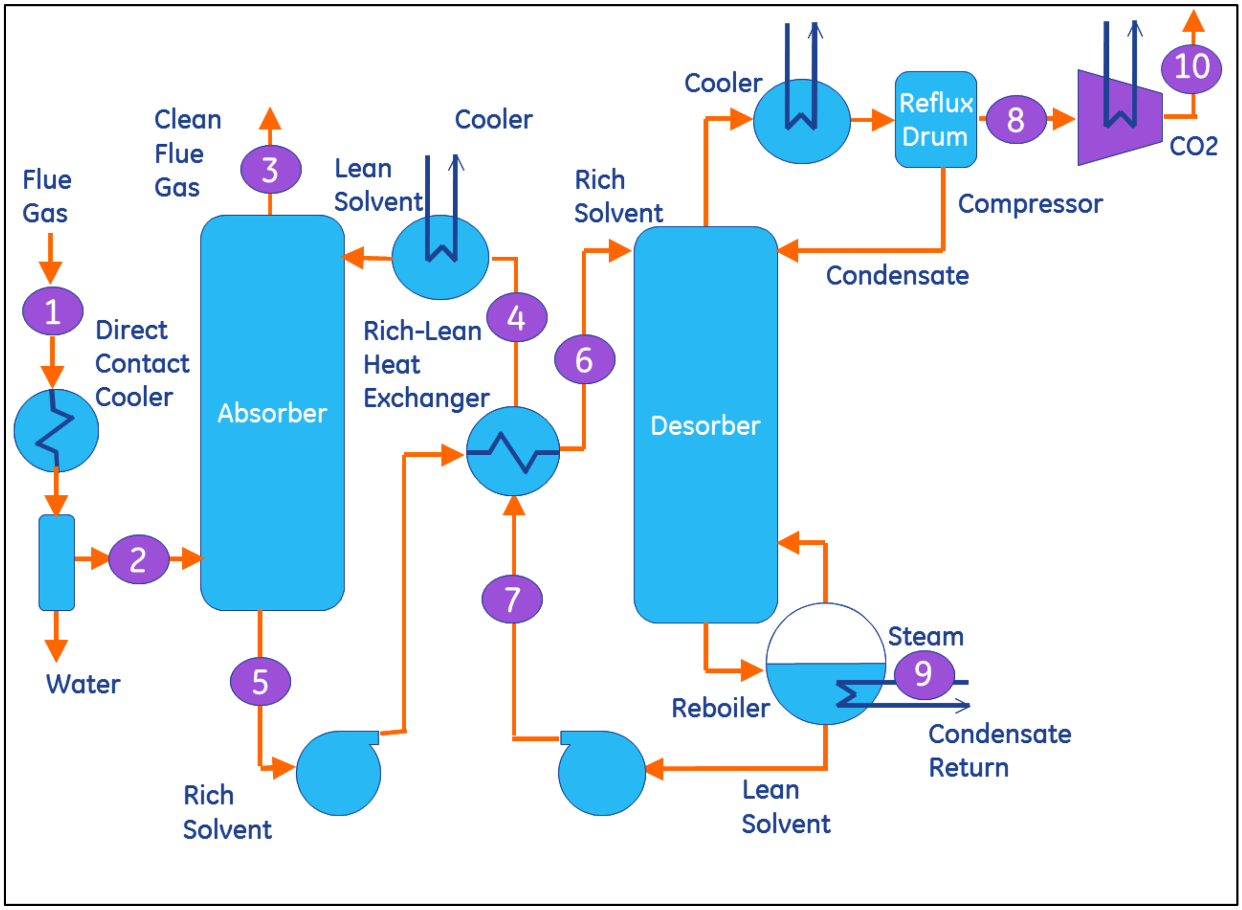


Figure 2. CO_2 separation sub-system.

For the amino-silicone solvent baseline case, the desorber operates at 140 °C (284 ^oF) and 4.3 atm (63 psia). For the sensitivity studies, the desorber conditions were varied from 130 to 140 °C (266 to 284 °F) and from 1.4 to 5.1 atm (20 to 63 psia) and these results are presented in subsequent sections. For the MEA Baseline Case, the desorber reboiler conditions are about 116 °C (240 °F) and 1.6 atm (23 psia). Steam is supplied to the desorber to provide heat, which releases CO_2 from the rich sorbent. Steam is supplied from the low pressure (LP) section of the steam turbine in the power plant sub-system. Steam conditions were given in the cooperative agreement. The hot vapor from the top of the desorber consisting primarily of CO_2 is cooled in a heat exchanger utilizing water. The stream then flows to a separator where the vapor and entrained liquid are separated. The CO_2 gas is removed from the separator and then delivered to the CO_2 product compressor. The liquid from the bottom of the separator is returned back to the desorber.

The lean sorbent from the desorber is pumped through the rich-lean heat exchanger to the absorber. The lean sorbent is cooled further before being fed to the absorber in order to increase the loading of CO_2 in the absorber.

GAP-1/TEG Plant-Scale Model Development

The absorber and desorber models developed in the first quarter of 2013 were combined to build an Aspen Plus^{TM} process model of the bench-scale process. The combined model is shown in Figure 3 below.

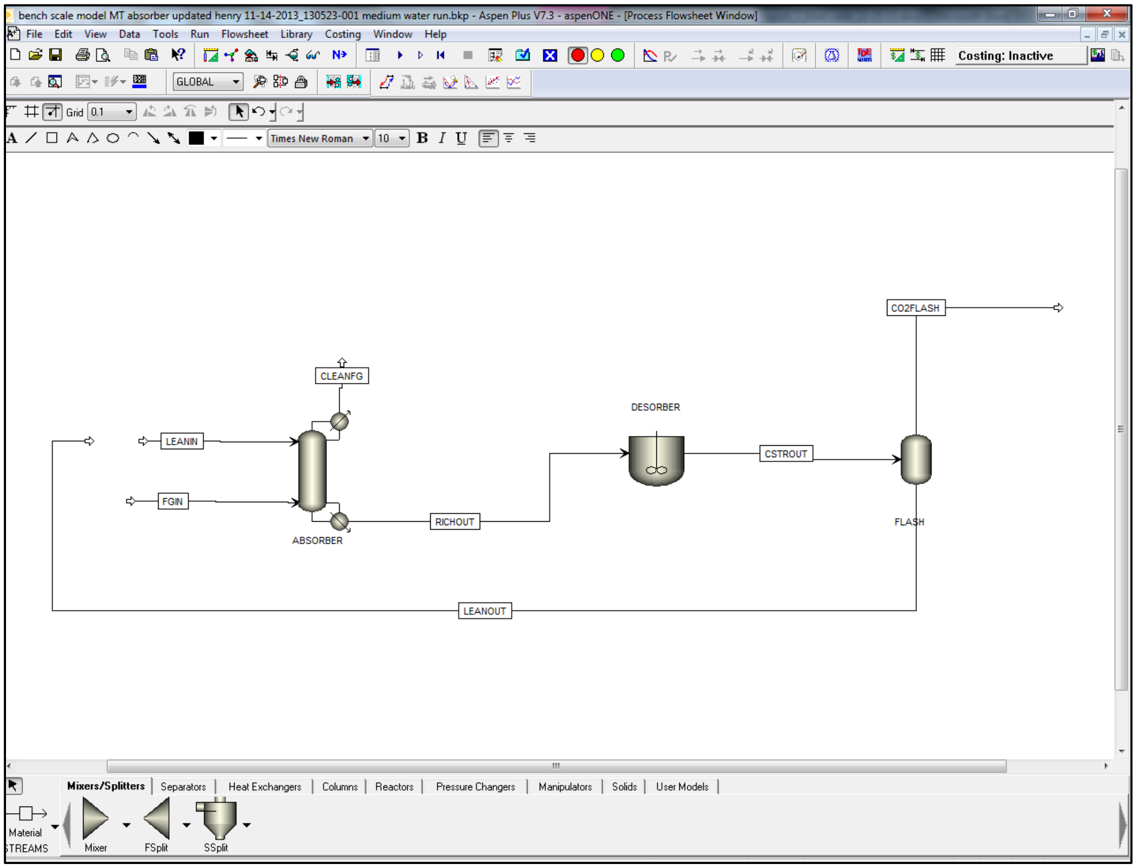


Figure 3. Bench-scale model of GAP-1/TEG process.

One of the biggest challenges in model development was a proper physical properties set-up because GAP-1 is a novel compound. Aspen Plus^{TM} does not have built-in properties for it. Molecular structure, molecular weight, and boiling point were manually input into the model. Also, some properties such as vapor pressure and viscosity were measured experimentally and then regressed in the model. Figure 4 compares experimental vapor pressure of GAP-1 with the model.

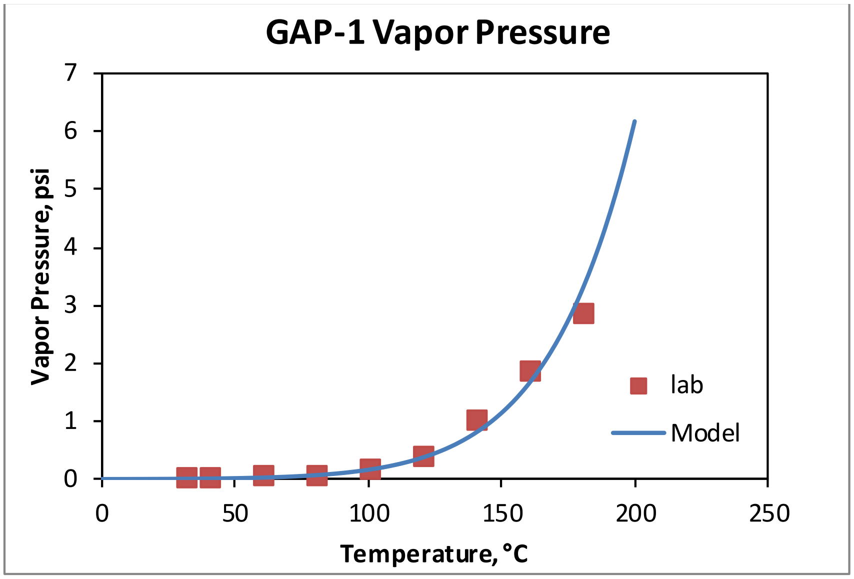


Figure 4. Comparison of experimental values of vapor pressure of GAP-1 with model values.

Henry's law constants for CO_2 in TEG, GAP-1 and GAP-1 carbmate are important values, but the Aspen Plus^{TM} Database does not have them built-in as well. For the values of Henry's law constants for CO_2 in TEG, ASPEN HYSYS software was used. It has a glycol properties package which is widely used in the oil and gas industries for modeling of the process to dry natural gas with TEG. The simple model, which was developed in HYSYS, is presented below in Figure 5.

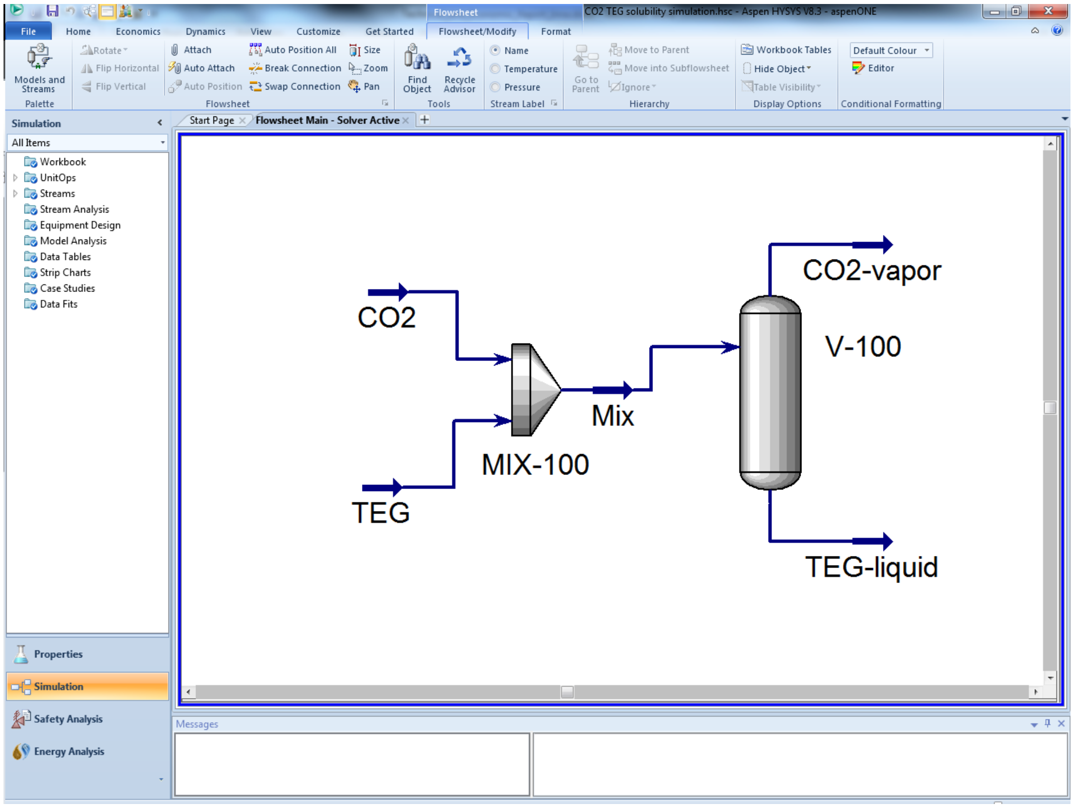


Figure 5. ASPEN HYSYS model to determine solubility of CO_2 in TEG.

An array of temperatures from 25 to 75 °C, and pressures from 200 to 500 kPa was selected to determine the solubility of CO_2 in TEG at equilibrium conditions, and the results are presented below in Figure 6.

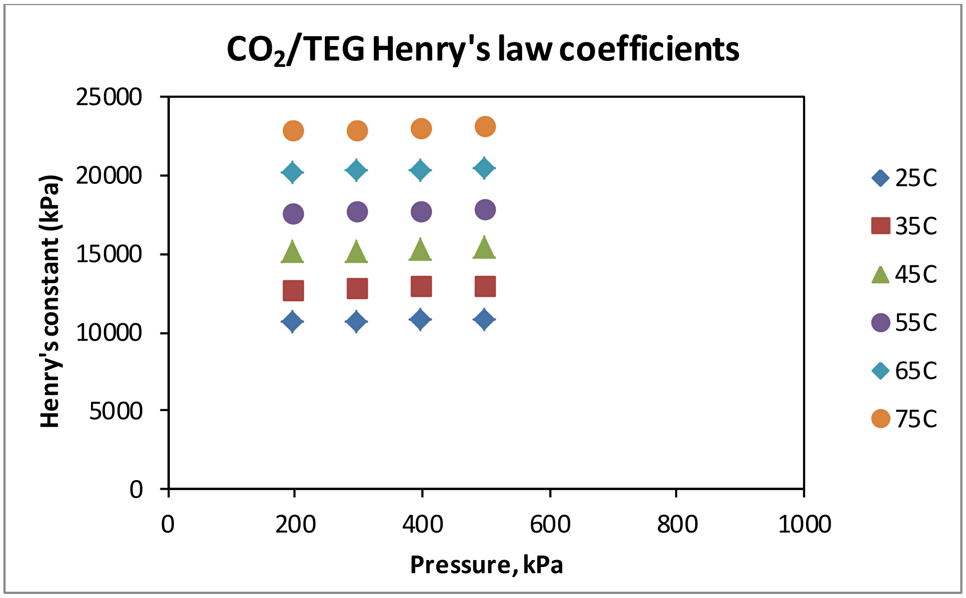
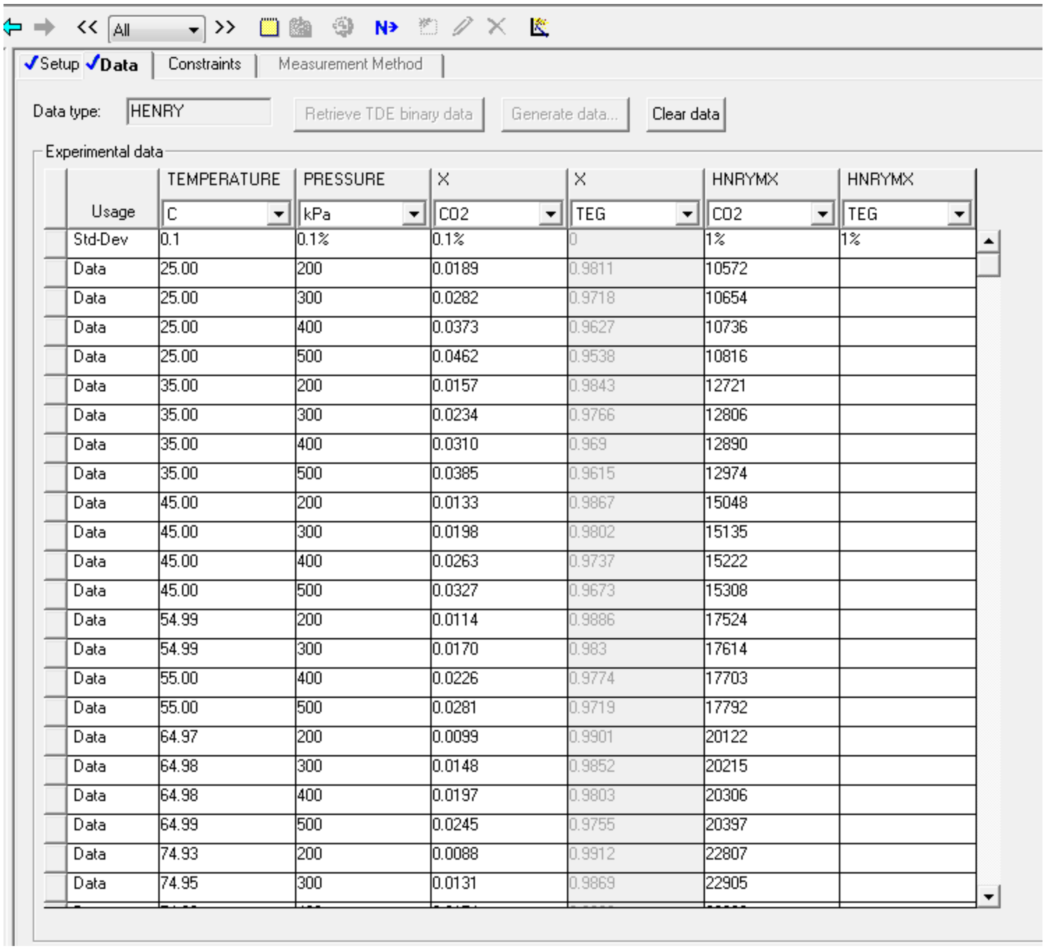


Figure 6. Henry's law constants of CO_2 in TEG based on ASPEN HYSYS modeling.

These values were used to calculate Henry's law constants and they were input into the DATA information of Aspen Plus^{TM} for regression as it shows in Figure 7.



Data type: HENRY

Usage	TEMPERATURE C	PRESSURE kPa	X CO2	X TEG	HNRYMX CO2	HNRYMX TEG
Std-Dev	0.1	0.1%	0.1%	0	1%	1%
Data	25.00	200	0.0189	0.9811	10572	
Data	25.00	300	0.0282	0.9718	10654	
Data	25.00	400	0.0373	0.9627	10736	
Data	25.00	500	0.0462	0.9538	10816	
Data	35.00	200	0.0157	0.9843	12721	
Data	35.00	300	0.0234	0.9766	12806	
Data	35.00	400	0.0310	0.969	12890	
Data	35.00	500	0.0385	0.9615	12974	
Data	45.00	200	0.0133	0.9867	15048	
Data	45.00	300	0.0198	0.9802	15135	
Data	45.00	400	0.0263	0.9737	15222	
Data	45.00	500	0.0327	0.9673	15308	
Data	54.99	200	0.0114	0.9886	17524	
Data	54.99	300	0.0170	0.983	17614	
Data	55.00	400	0.0226	0.9774	17703	
Data	55.00	500	0.0281	0.9719	17792	
Data	64.97	200	0.0099	0.9901	20122	
Data	64.98	300	0.0148	0.9852	20215	
Data	64.98	400	0.0197	0.9803	20306	
Data	64.99	500	0.0245	0.9755	20397	
Data	74.93	200	0.0088	0.9912	22807	
Data	74.95	300	0.0131	0.9869	22905	

Figure 7. Henry's law constants of CO_2 in TEG in DATA tab in Aspen Plus^{TM}.

Regression fit of these data is shown on Figure 8.

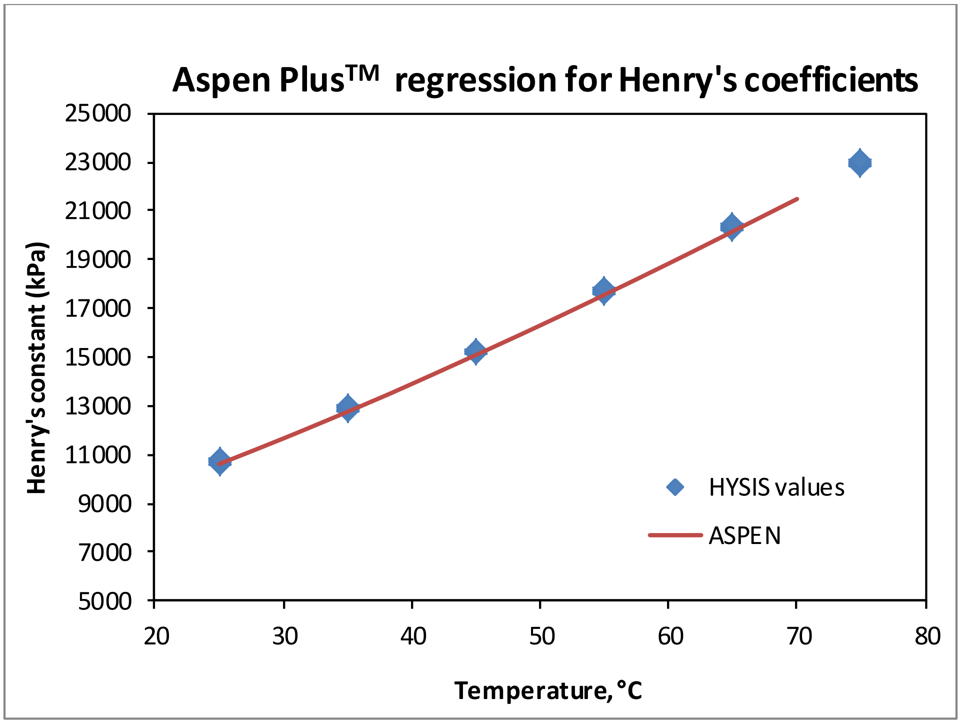


Figure 8 Regression of Henry's law constants of CO_2 in TEG in Aspen Plus^{TM}.

It was challenging to measure the solubility of CO_2 in GAP-1 experimentally, because of the chemical reaction of CO_2 with the GAP-1 and the inability to separate physical solubility and reaction. It was also challenging to experimentally measure solubility of CO_2 in GAP-1 carbamate. The Aspen Plus^{TM} database has the properties for D4, octamethylcyclotetrasiloxane, and it was assumed that the solubility of CO_2 in GAP-1 and GAP-1 carbamate are similar to the solubility in D4.

A set of bench-scale steady-state experiments was selected using an experimental design of experiments (DOE) that was completed during the second quarter of 2013. The experiments were selected such that they were representative of the overall DOE matrix for absorber and desorber conditions. Five experiments were selected to validate the desorber model predictions with experimental results. The process conditions for these five experiments are shown in Table 1. The main parameters that were changed were the temperature and the pressure of the desorber. The inlet gas-phase CO_2 concentration to the absorber was around 16% in all the experiments.

Table 1. Five experiments selected to confirm bench-scale model performance.

Case	**Temperature (°C)**	**Pressure (psig)**	**Actual CO_2 % inlet to absorber**
#1:130522-292	140	0	16.54%
#2:130523-001	140	45	15.99%
#3:130530-021	140	45	15.72%
#4:130606-058	150	45	15.92%
#5:130618-082	120	45	16.24%

The liquid solvent composition at the exit of the desorber (inlet of the absorber) is a critical parameter to observe for this set of experiments. The model predictions were compared with the experimental results. There was good agreement between the predicted and the measured values as shown in Figure 9. Also, the effect of increasing pressure and decreasing temperature is observed in both experimental results and model predictions. Comparison of case #1 with Case #2 and Case #3 shows the effect of increasing pressure. With increase in desorber pressure the GAP-1 concentration decreases and GAP-1-carbamate concentration increases. Comparison of Case #2 and Case #3 with Case #4 and Case #5 shows the effect of desorber temperature. With increase in desorber temperature the GAP-1 concentration increases and GAP-1-carbamate concentration decreases.

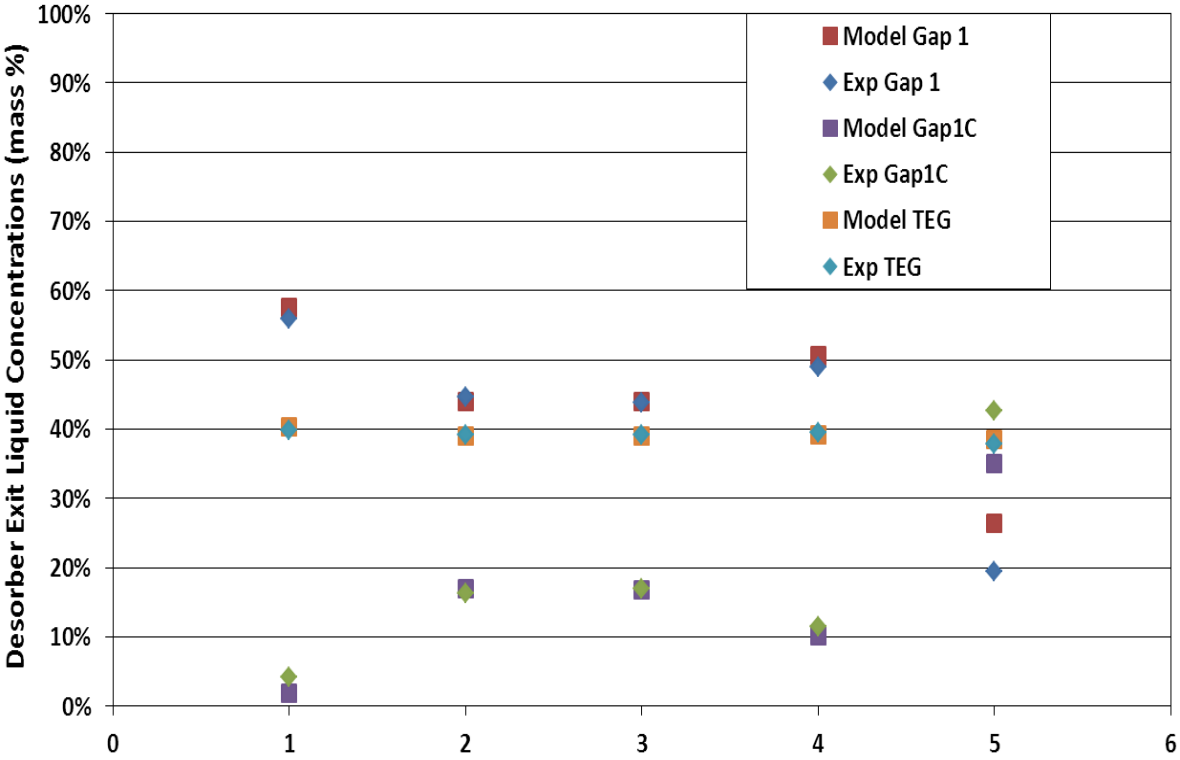


Figure 9. Comparison of experimental results and model predictions for desorber outlet concentrations of GAP-1, TEG, and GAP-1-carbamate.

The absorber response was also compared for these five experiments. The liquid and gas flow rates into the absorber were not varied between these five experiments. The desorber exit liquid was recycled back into the absorber after cooling the lean solvent.

The absorber was modeled by two separate methods. In the first method the mass transfer and the chemical reactions taking place were modeled as equilibrium (top two charts in Figures 10-14) whereas in the second method the mass transfer was modeled as rate and the reaction was modeled as equilibrium (bottom two charts in Figures 10-14).

The absorber response observed experimentally was compared with the model predictions. Two main parameters were selected for the comparison, the gas phase CO_2 concentration profile and temperature profile along the length of the absorber. The next set of figures show these comparisons (Figures 10-14). It is observed that the second method, using the rate/equilibrium model, gives very good predicted values when compared to the experimental observations for both the CO_2 concentration as well as the temperature profile in the absorber. The equilibrium/equilibrium model is less consistent with experimental observation.

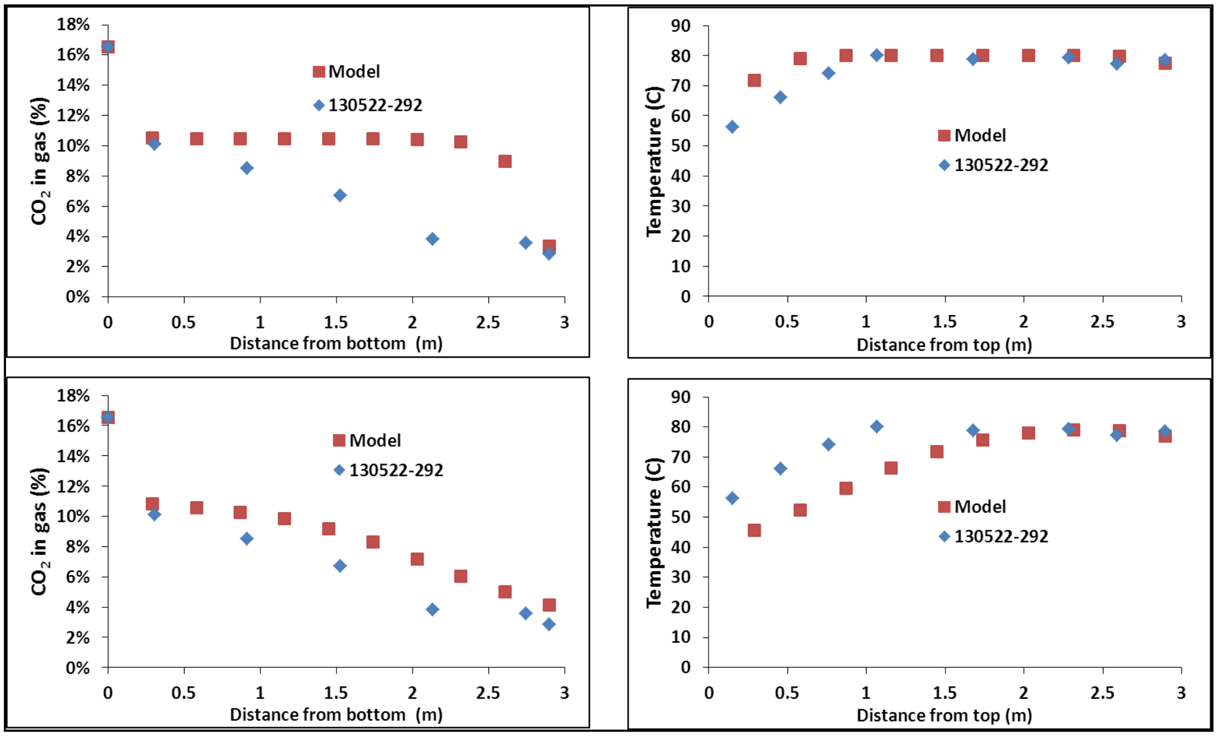


Figure 10. Absorber response for Case #1. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

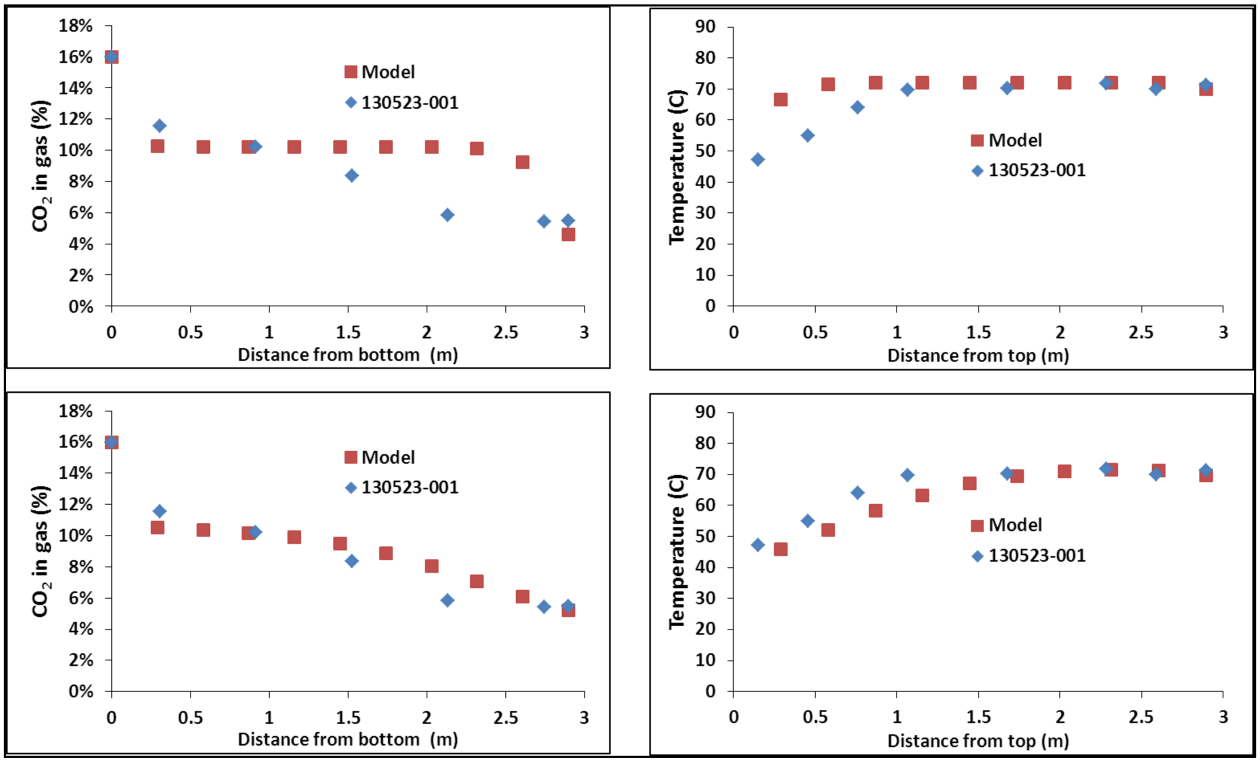


Figure 11. Absorber response for Case #2. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

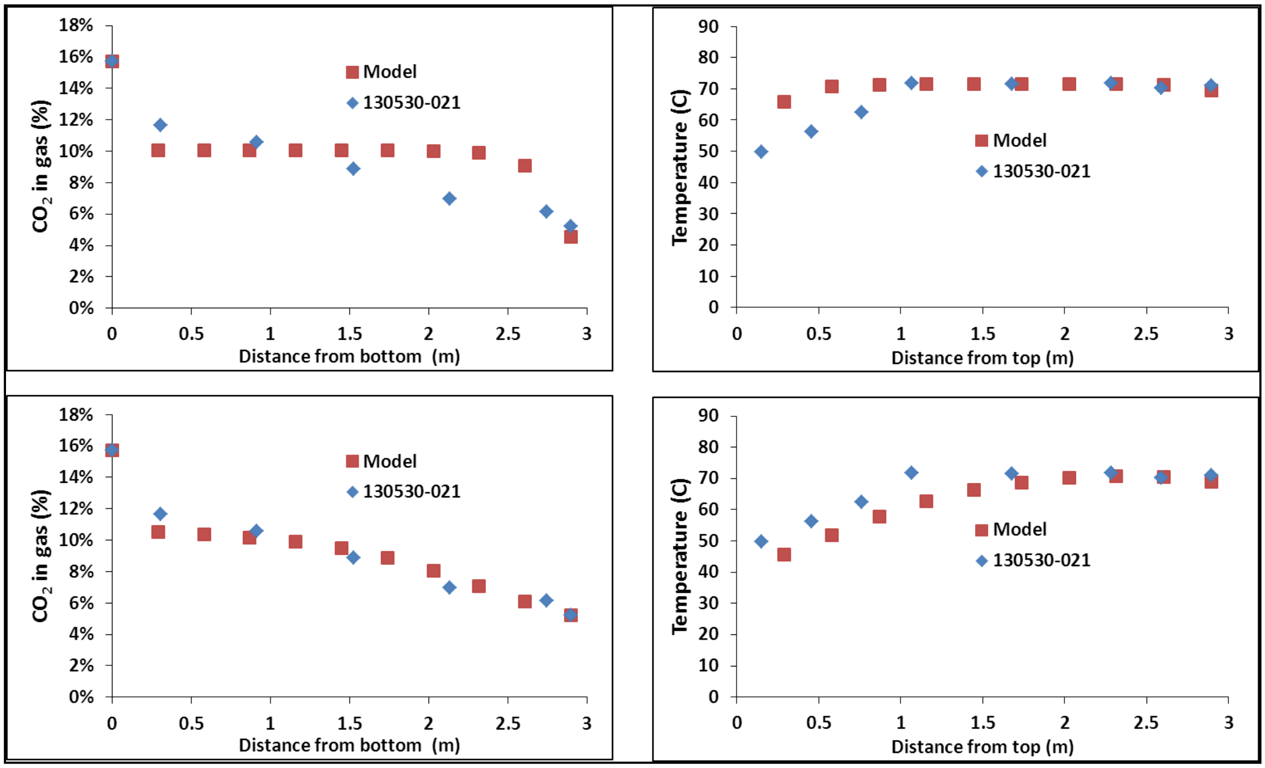


Figure 12. Absorber response for Case 3. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

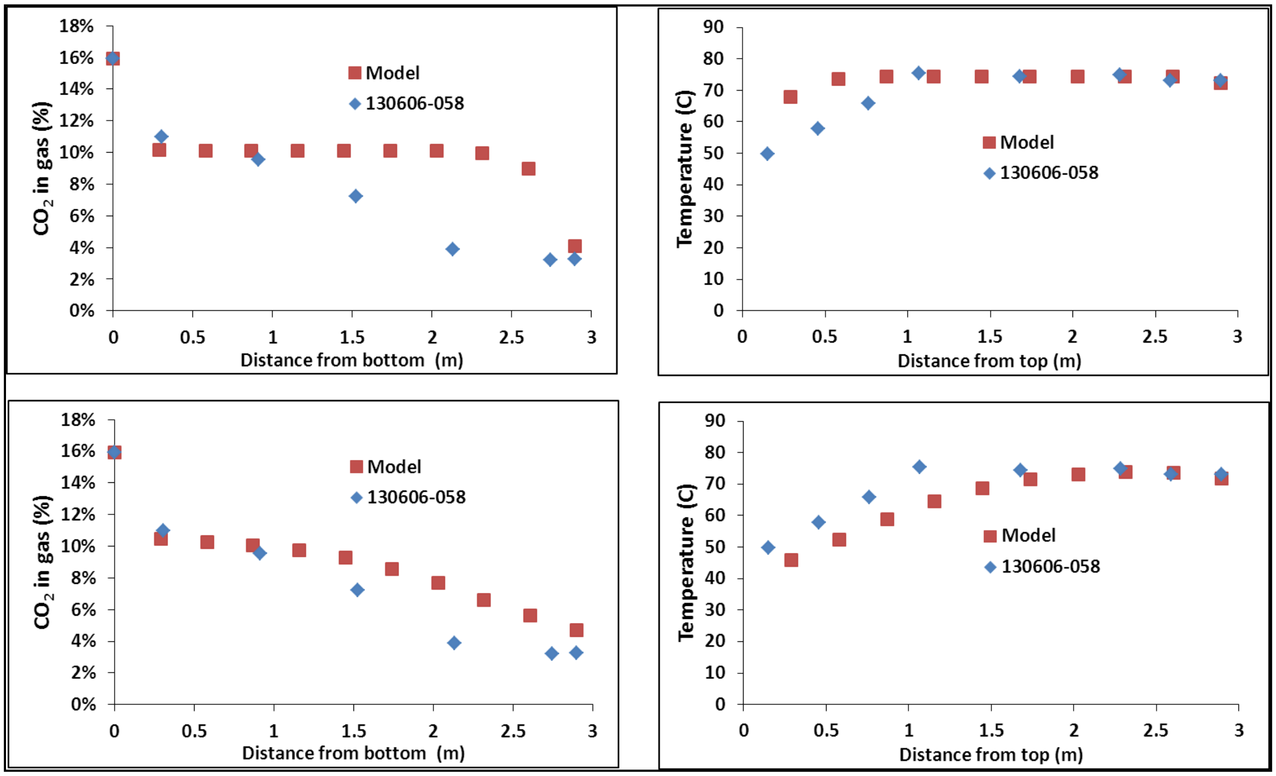


Figure 13. Absorber response for Case #4. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

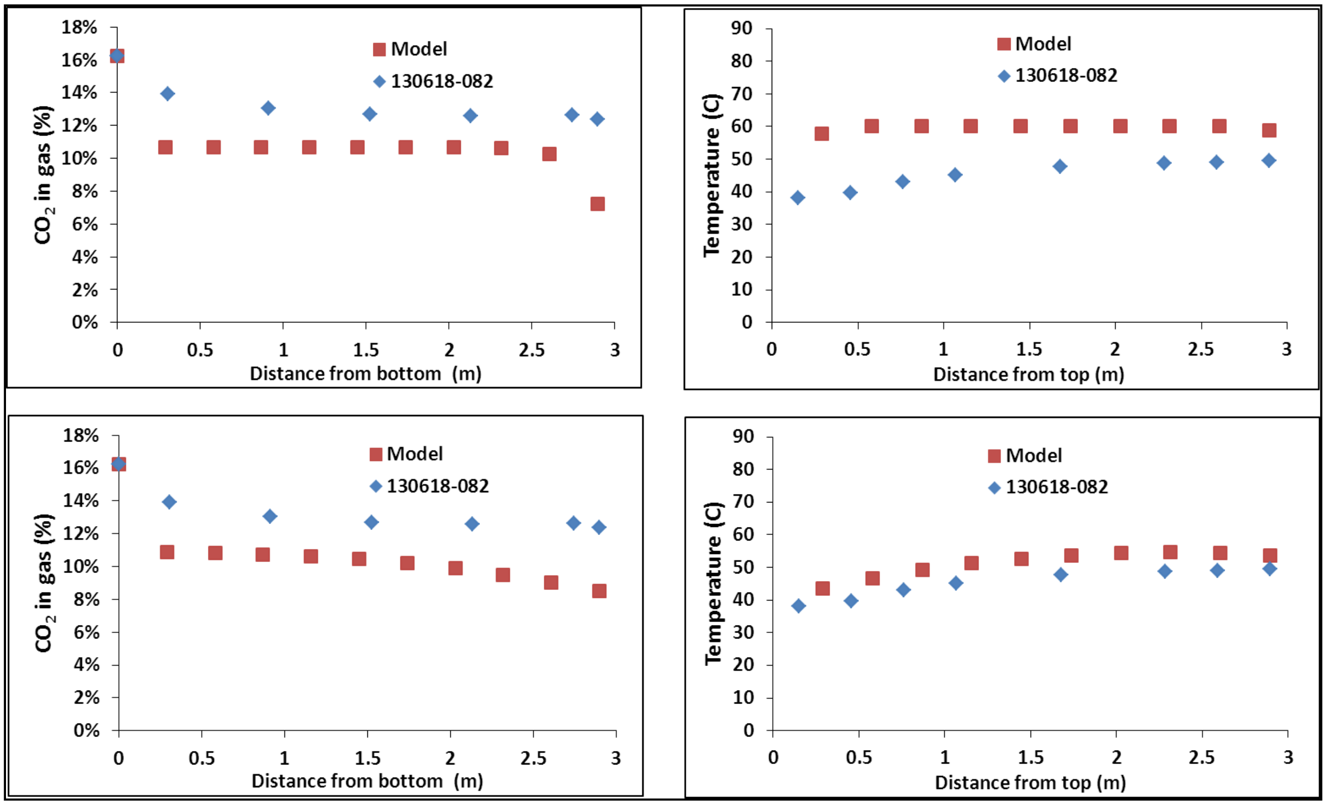


Figure 14. Absorber response for Case #5. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

Four more experiments were selected to compare the absorber response with the model predictions. In these selected experiments the gas flow rate and the liquid (lean solvent) flow rate were varied as shown in the Table 2 below. The CO_2 concentration in the gas phase inlet to the absorber was fixed around 16%. Note that the temperature profile for Case #6 was not available experimentally. Good agreement was again observed between the experimental results and the model predictions via the second method, the rate/equilibrium model. Generally, excellent agreement was observed at the inlet and outlet conditions. The concentration and temperature profiles showed generally good agreement even though there were some differences between the rate/equilibrium model and the experimental observations in some of the selected cases. The comparisons are shown in Figures 15-18.

Table 2. Experiments selected for comparison of model predictions with experimental observations.

Case	**Gas flow rate (slpm)**	**Liquid flow rate (lpm)**	**Actual CO_2 % inlet to absorber**
#6:130411-152	**75**	**0.5**	**16.10%**
#7:130422-167	**100**	**0.5**	**16.27%**
#8:130423-177	**119**	**0.5**	**16.29%**
#9:130509-257	**119**	**1**	**16.14%**

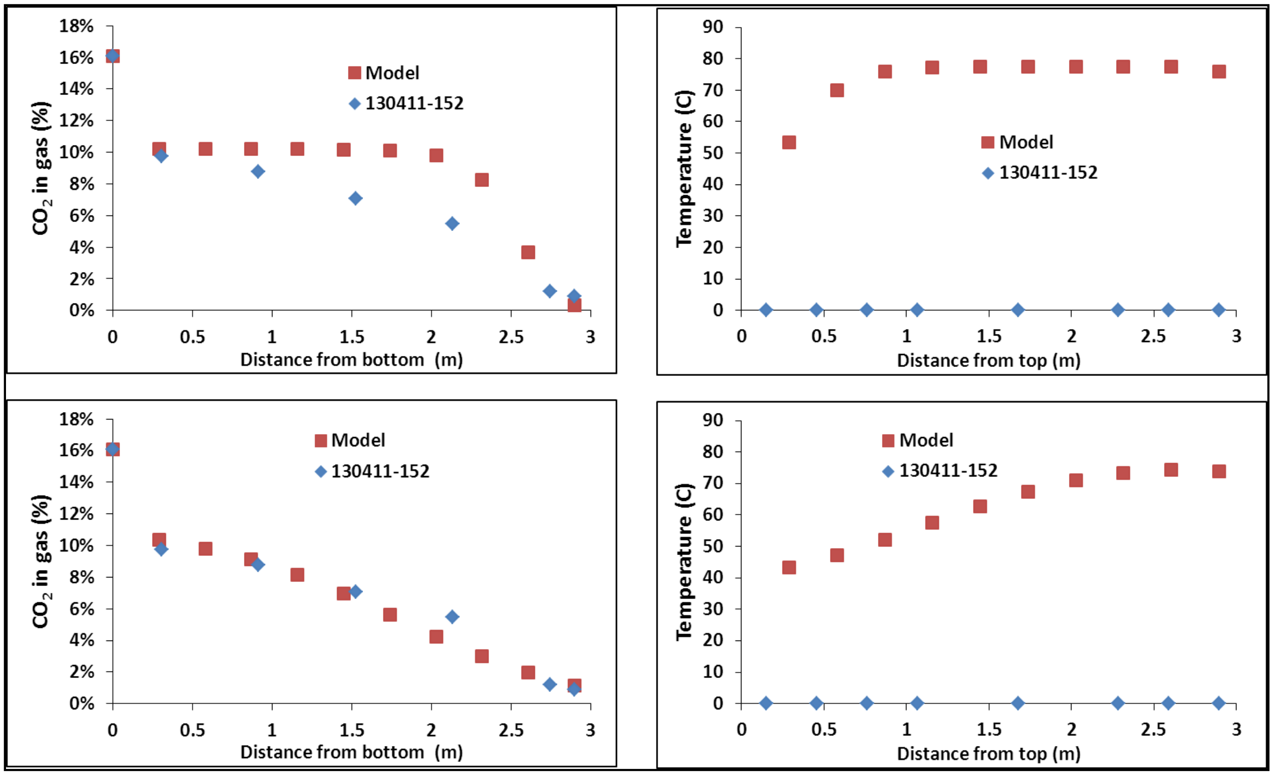


Figure 15. Absorber response for Case #6. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model. The experimental temperature profile was not measured in this experiment.

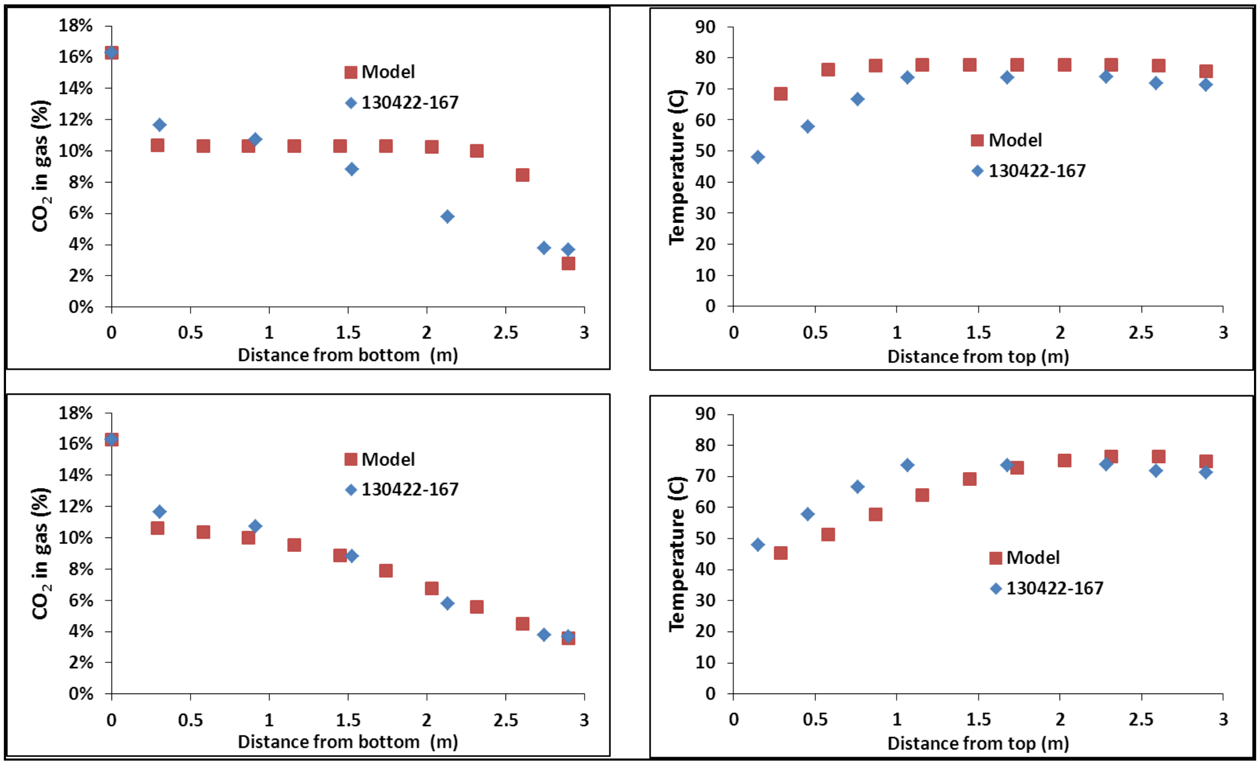


Figure 16. Absorber response for Case #7. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

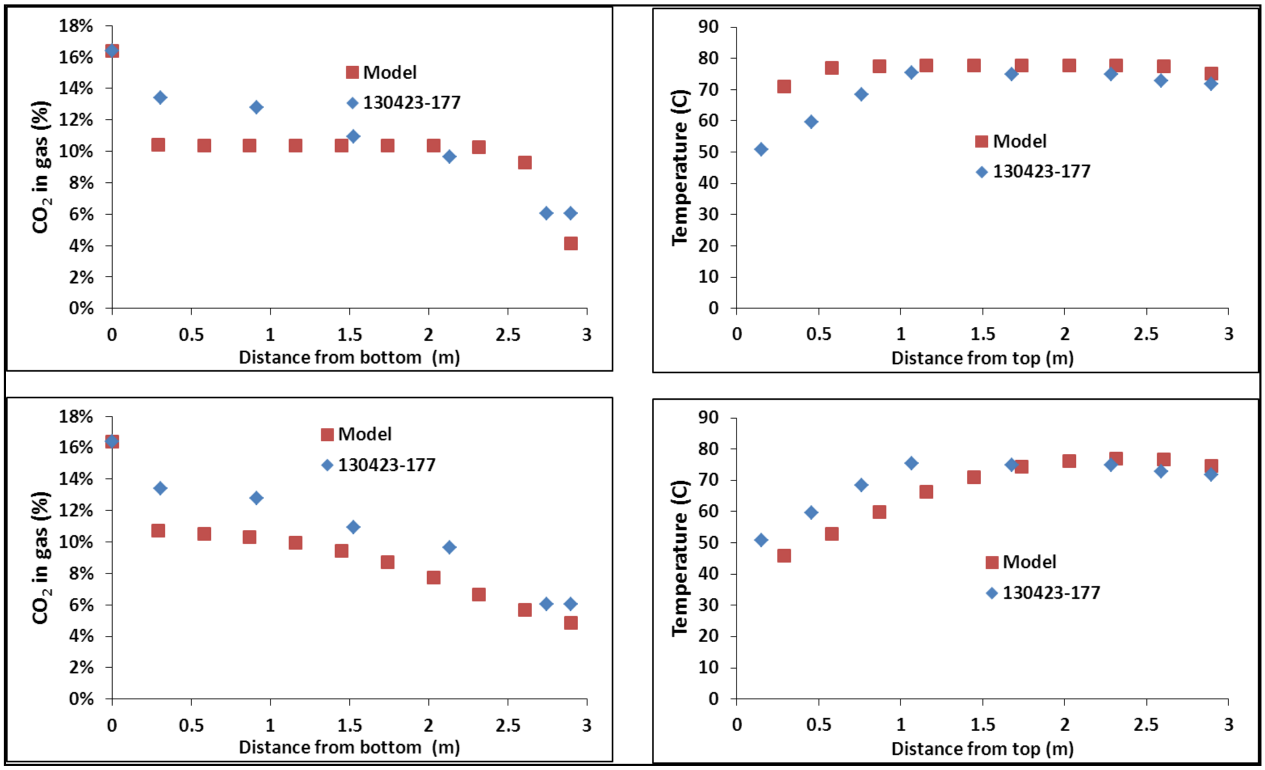


Figure 17. Absorber response for Case #8. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

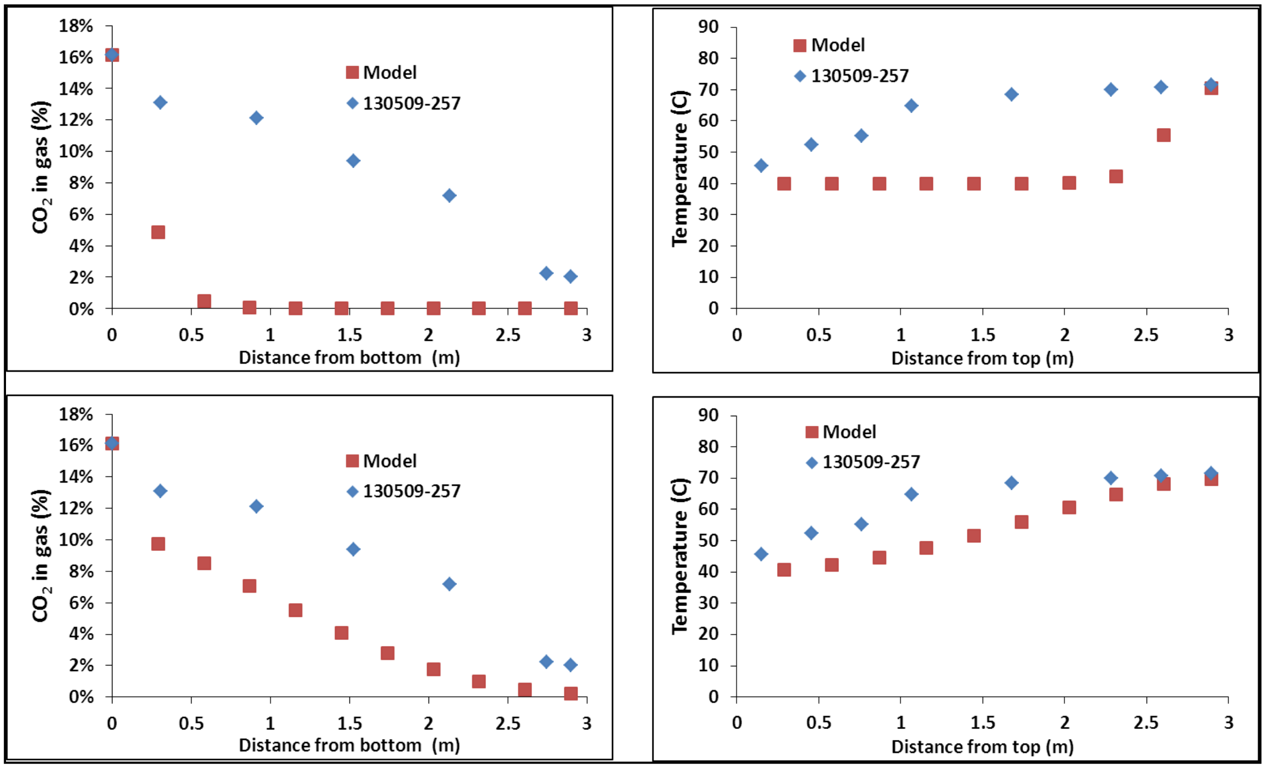


Figure 18. Absorber response for Case #9. The top two charts use the equilibrium/equilibrium model in Aspen Plus^{TM}. The bottom two charts use the rate/equilibrium model.

In one of the cases for economic evaluation, the temperature of the desorber was taken to be 130 °C, and therefore additional comparison of three bench-scale experiments and model predictions was conducted. These runs were at the following conditions.

Table 3. Operating conditions for runs compared with bench-scale model performance.

	Run 130605-047	**Run 130620-100**	**Run 131003-060**
Desorber T,°C	130.8	131.5	130
Desorber P, psig	45.3	45.3	45
%mol CO_2 in FG	16.01%	15.94%	16.80%
Gas flowrate, SLPM	112	112	119
Liquid flowrate, lpm	0.5	0.5	1.8

These input data were properly set up in the Aspen Plus^{TM} bench-scale model, and for all analyses the absorber was modeled with Rate-based mass transfer since it demonstrated much better agreement with experiment. The results are presented below for each of the models. Error bars were determined through a Gage R analysis for the experimental data.

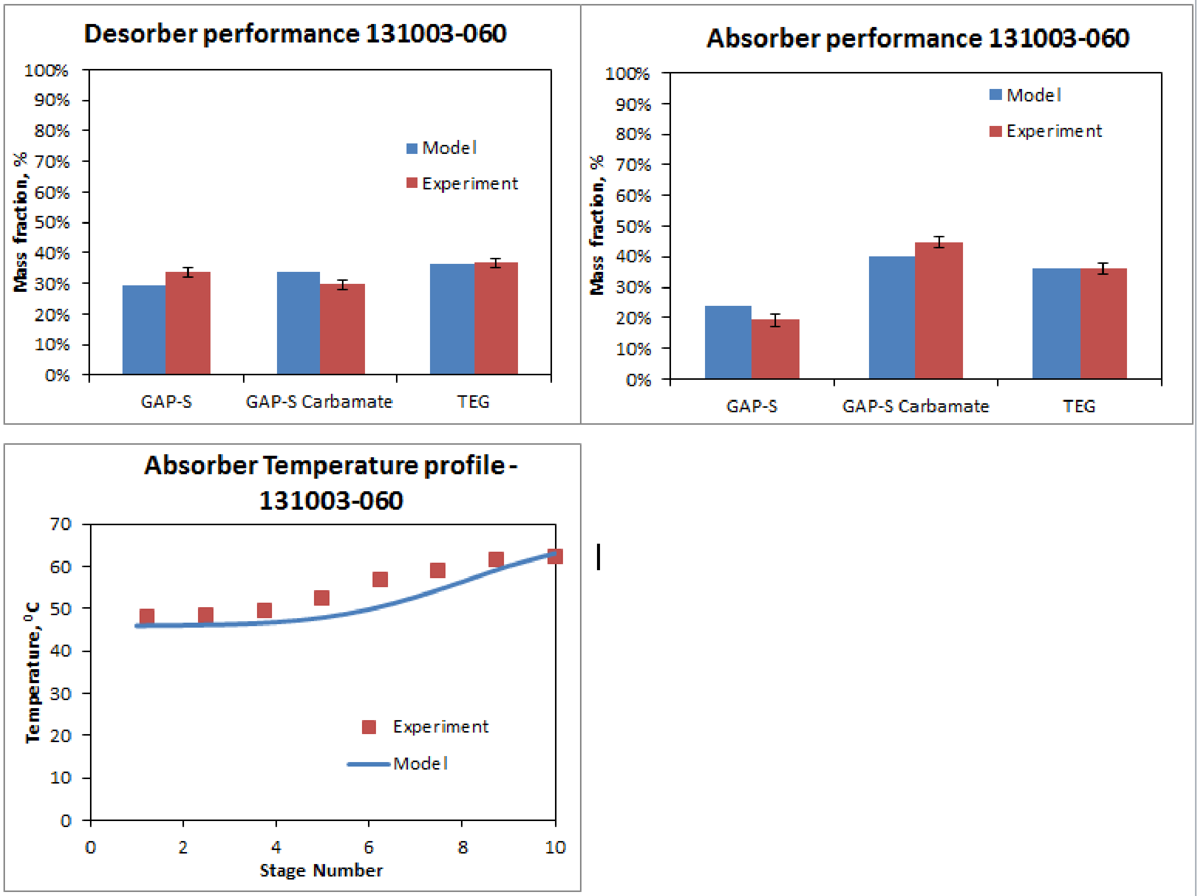


Figure 19. Comparison of bench-scale model with experimental data for Run 131003-060.

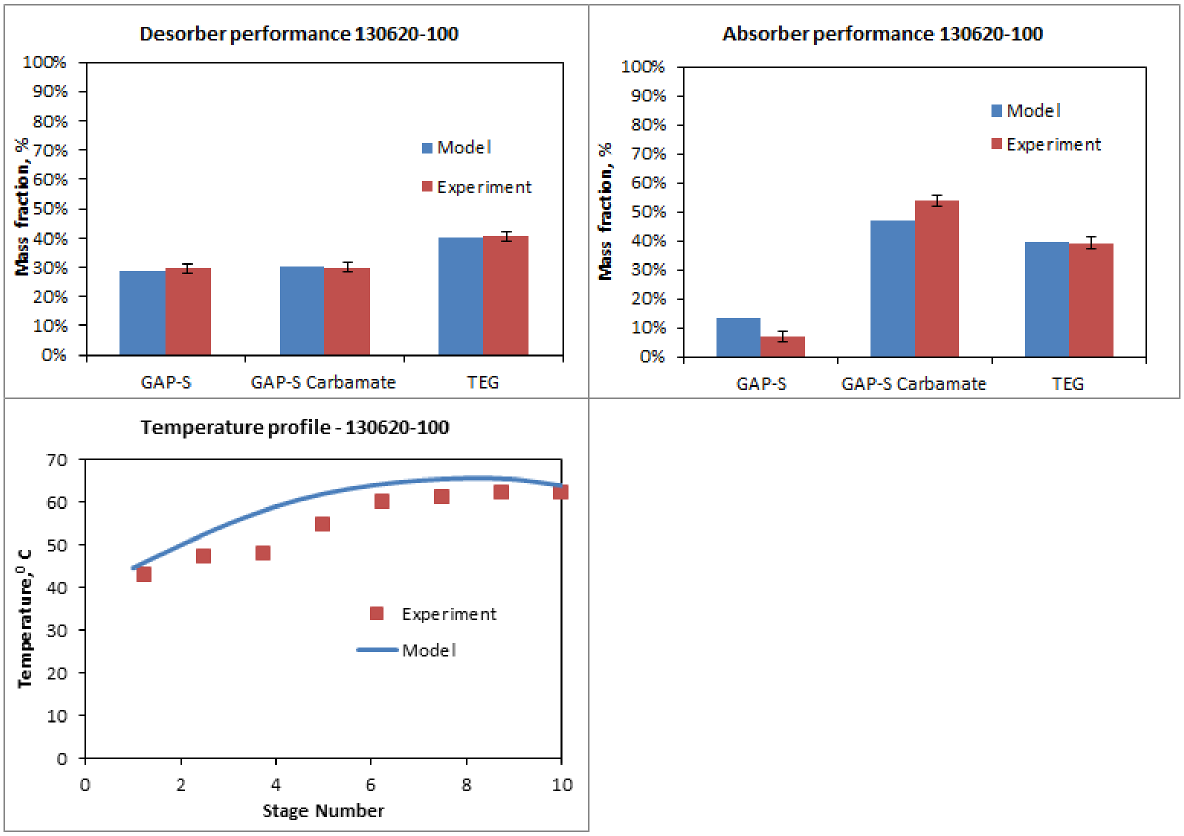


Figure 20. Comparison of bench-scale model with experimental data for Run 130620-100.

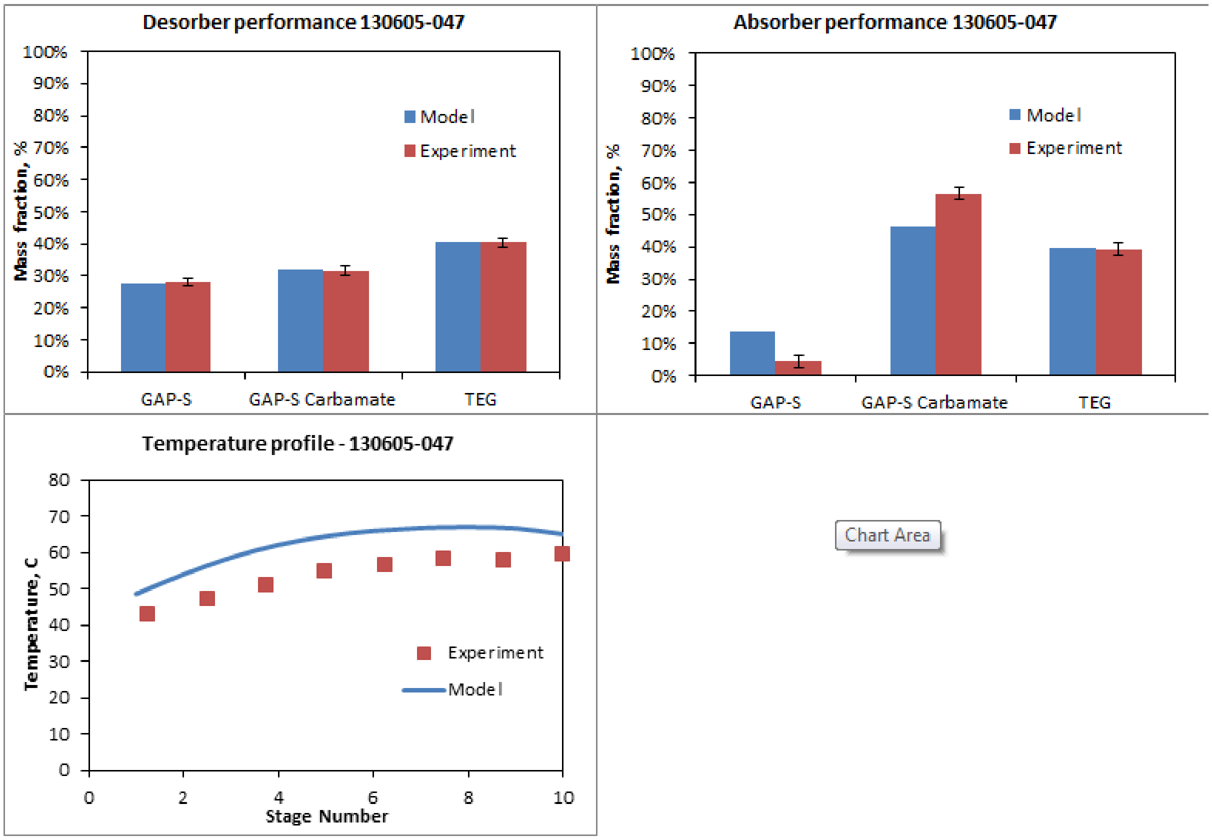


Figure 21. Comparison of bench-scale model with experimental data for Run 130605-047.

Since the bench-scale model demonstrated good predictability of the experimental data, it was used to build a more detailed model for the plant-scale GAP-1/TEG process, and it is presented in Figure 22.

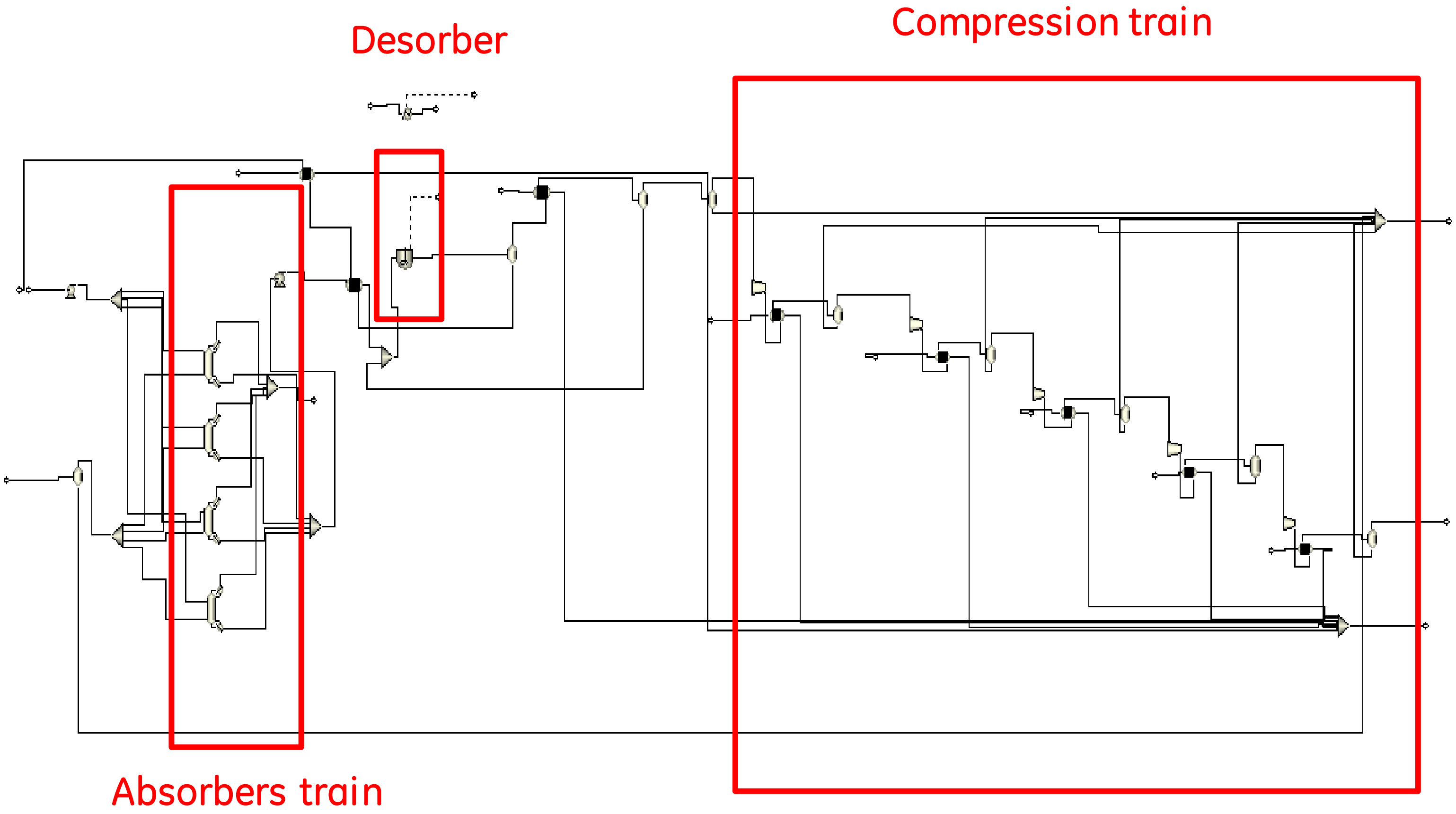


Figure 22. Plant-scale Aspen Plus^{TM} model for 60/40 GAP-1/TEG CO_2 capture process.

Each step of this model will be discussed in detail.

Absorber

Absorber train consists of 4 absorbers and the flow diagram is presented below.

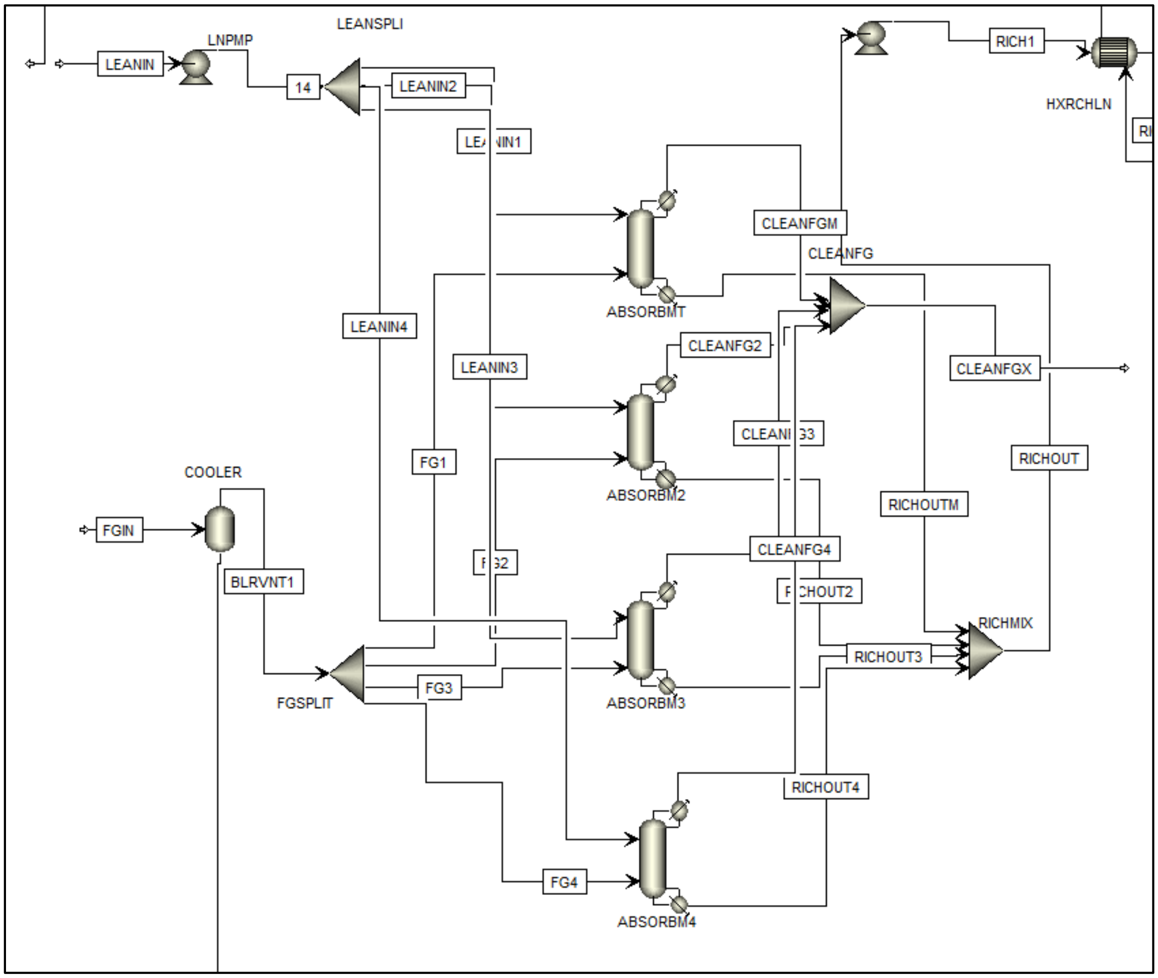


Figure 23. Absorber train of 4 absorbers for 60/40 GAP-1/TEG CO_2 capture plant-scale process.

Significant improvements in the absorber model have been made this year. In the preliminary Technical and Economic Feasibility Study, the absorber was modeled as a yield reactor. In the current model each absorber is modeled using the RadFrac block which models packed towers with chemical reaction. The same set-up is used for the absorbers in the detailed model for the MEA processes.

Packing information is built into the Aspen Plus^{TM} software for specific types of packing. The packing most similar to the packing in the bench-scale model was used. The diameter of the absorber column is determined mainly by the gas flow rate, and an iterative analysis was conducted to determine minimal diameter without flooding in the column (38 ft).

Height of the packing in the absorber is determined by 2 main factors, mole ratio of total liquid flow rate (Lm) to the total gas flow rate (Gm) and the required CO_2 capture by the absorber, which is 90%. Sensitivity analysis in the model was conducted to determine the minimal packing height required to obtain 90% of CO_2 capture at fixed Lm/Gm ratio. From the results of this sensitivity analysis, a packing height of 50 ft was selected, with a slight increase in lean solvent flow rate required to obtain the specified CO_2 capture.

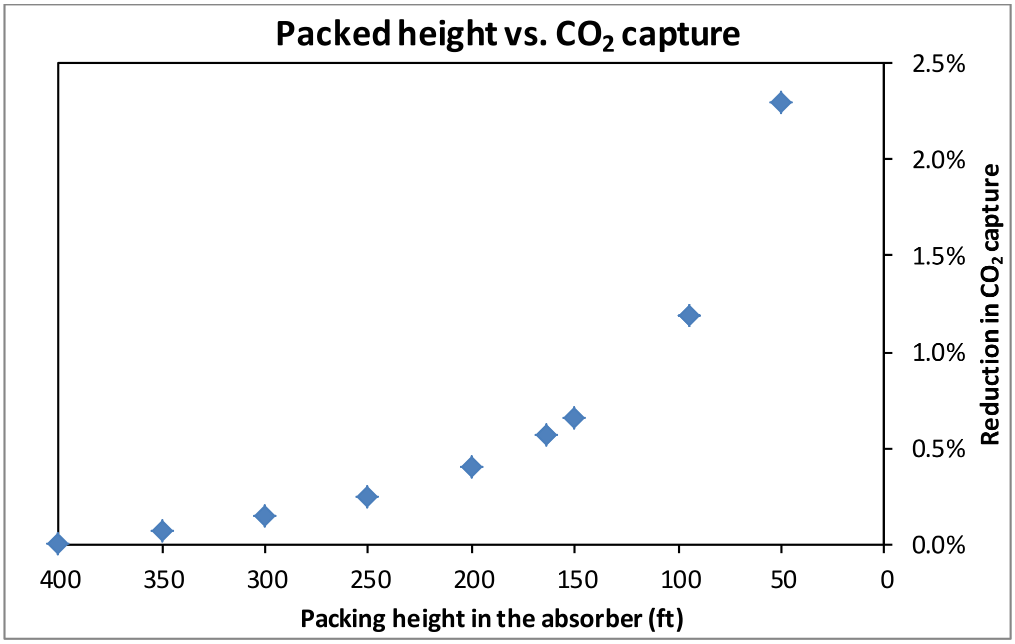


Figure 24. Reduction in CO_2 capture versus height of packing in absorber relative to 400 ft of packing.

In this Technical and Economic Feasibility Study several cases were considered to determine the optimal economic option. The following cases were considered.

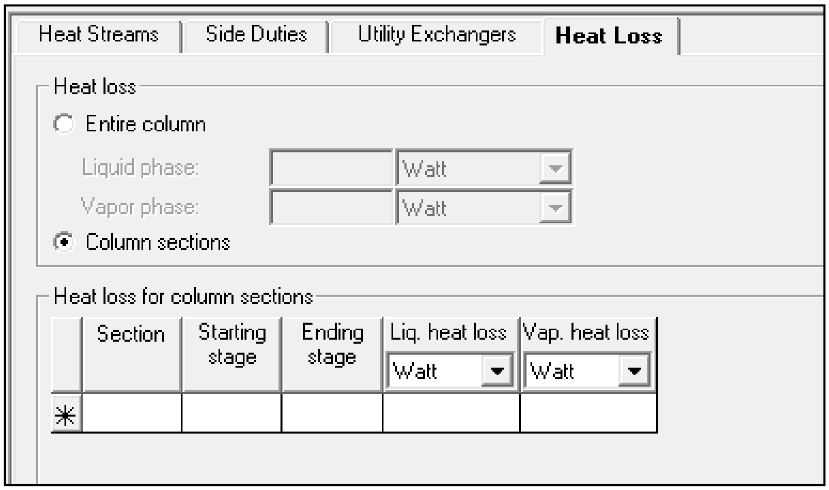
Table 4. Process conditions for cases considered for economic analysis and comparison.

	Lm/Gm	*Desorber T, °C*	*Desorber P, psia*	*Intercoolers*	*Material*
Case 1	1.12	140	63	No	CS/SS
Case 2	0.81	140	63	Yes	CS/SS
Case 3	0.66	140	20	Yes	CS/SS
Case 4	0.72	130	20	Yes	CS/SS
Case 6	0.99	130	63	Yes	CS/SS
Case 7	0.81	140	63	Yes	CS

In each case 90% CO_2 capture was obtained. Since the flue gas flow rate is fixed for all of the cases, the dimensions of the absorber units are the same. The diameter of the absorber is 38 ft, and the packed height is 50 ft.

Case 1 primarily uses carbon steel (CS) components with stainless steel (SS) used in the most corrosive portions of the process. Case 2 and higher have absorber stage intercoolers incorporated into the model. They reduce the temperature of the solvent in the absorber. By doing so, more CO_2 is captured per lean solvent amount, therefore reducing the Lm/Gm ratio. There are several ways to set up intercoolers in Aspen Plus^{TM} software. It can be set up as a total heat load removed or amount removed per stage from the vapor and liquid phases. Case 7 uses carbon steel for all components.

During bench scale experiments some heat was lost into the environment, although the bench-scale model doesn't incorporate this and considers the absorber to perform adiabatically. Some iterative work was conducted to determine how temperature profile in the model is affected by heat loss from the solvent. One experimental run was selected, and the temperature profile was analyzed at different levels of heat removal.



Heat loss

○ Entire column

Liquid phase: Watt

Vapor phase: Watt

● Column sections

Heat loss for column sections

	Section	Starting stage	Ending stage	Liq. heat loss Watt	Vap. heat loss Watt
*					

Figure 25. Heat loss set-up for the absorber in Aspen Plus^{TM}.

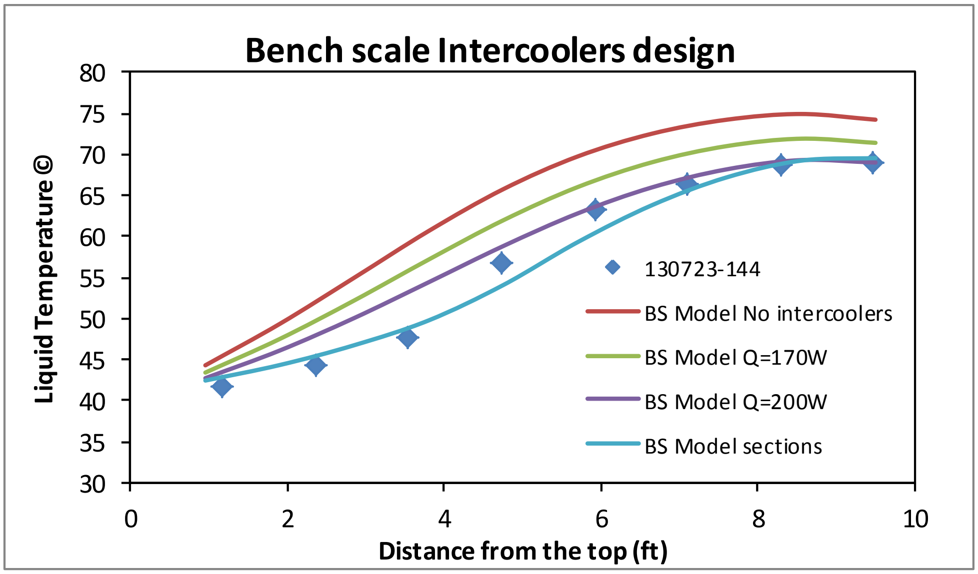


Figure 26. Effect of different heat removal amounts on temperature profile in the absorber.

It can be see that by addition of the intercoolers into the model the model temperature profile better matches the experimental bench-scale data. During the following pilot-scale program the absorber will be equipped with intercoolers. Therefore, the amount of heat removed from the column will be known and will be incorporated in the process model.

Desorber Design

The desorber was designed as a CSTR for the bench-scale system, and it will be scaled up as a CSTR as well. The plant scale model for the desorber unit is presented below.

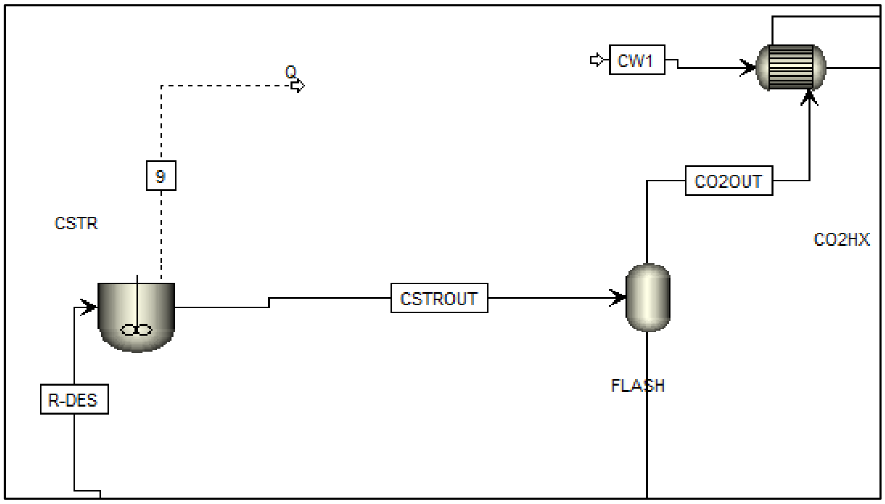


Figure 27. Desorber unit of plant-scale model.

The main design parameters for the CSTR are temperature, pressure, and residence time. Temperature and pressure were varied for different cases (see above). Residence time was selected to be 11 minutes because a sensitivity analysis determined that to be the minimum time at which the process reaches equilibrium.

Since the desorber unit is a critical unit operation in this process, redundacy will be required at the commercial scale to ensure safe process operation. Therefore, the desorber system consists of two CSTR reactors, each with a recirculation loop. The recirculation loops include shell-tube heat exchangers and pumps. The recirculation loops server two functions. First, they will transfer required heat to the solvent, and second they will increase mixing, thus reducing the mass transfer limitations. Since solvent flow rate varies in each considered case, the size of the desorber and recirculation loop heat exchanger will vary in each case.

The values of the overall heat-transfer coefficicents for desorber jacket and shell tube heat exchangers was found in the literature, and the following values were used.

Table 5. Values for overall heat transfer coefficients for jacketed vessel and shell and tube heat exchanger.

Type of heat transfer unit	*Overall heat transfer coefficient U (BTU/hr·ft^2·F)*
Jacketed vessels: steam to organics, SS wall, average	100
Shell and Tube heat exchanger: steam to light organics, average	185

Based on a specified residence time of 11 minutes, the volume of the reactor can be determined. The amount of heat which can be transfered through the jacket of the reactor can be determined through the following equation.

$$Q_{reactor} = U * A * LMTD$$

And the heat required to be transferred through the recirculation loop is calculated by the difference of the total heat load and $Q_{reactor}$. Based on the obtained value, the size of the heat exchanger and number of cycles is determined. Below is the summary table for the desorber design for different cases.

Table 6. Conditions for cases considered for economic analysis.

	Desorber diameter, m	*Desorber height, m (includes 5 m V-L disengagement volume)*	*Area of Recirculation loop heat exchanger, m^2*	*Number of cycles/min*
CASE 1	10	20	519	0.34
CASE 2	10	16	534	0.43
CASE 3	10	15.4	572	0.50
CASE 4	10	15.9	553	0.45
CASE 6	10	18.2	503	0.36

These dimensions were used in the economic calculations.

CO_2 Separation Unit Key System Assumptions

The model used the following process design assumptions given in cooperative agreement DE-FE0007502.

1) Composition of flue gas leaving FGD (wet basis):

	Volume %
CO_2	13.17
H_2O	17.25
N_2	66.44
O_2	2.34
Ar	0.8
	ppmv
SOx	42
NOx	74

2) Flow rate of flue gas leaving FGD (based on 550 MW net PC plant): 5,118,399 lb/hr

3) Pressure and temperature of flue gas leaving FGD: 14.7 psia and 135 °F

4) Conditions for LP steam available from power plant: 167.7 psia and 743.3 °F

5) Conditions for cooling water: feed = 60 °F, return = 80 °F with a minimum approach of 30 °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 amino-silicone solvent baseline models are based on the typical temperature-swing sorbent separation process. The systems have four process variables that dominate the performance with a given sorbent 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 sorbent

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

The sorbent rich loading is defined as the weight % of CO_2 in the rich sorbent leaving the absorber column. The sorbent lean loading is defined as the weight % of CO_2 in the lean sorbent leaving the desorber column. The sorbent net loading is defined as the difference between the rich loading and the lean loading and was obtained from bench-scale experiments for the GAP-1/TEG system.

A detailed MEA Aspen Plus^{TM} model that was built under this project was used for comparison.

The main features of the MEA model include an absorber, rich-lean heat exchanger, and a desorber. The same unit operations are important for the GAP-1/TEG system. The baseline MEA case is built from the description given in the Volume 1: Bituminous Coal and Natural Gas to Electricity" report by National Energy Technology Laboratory.

Heat and material balances for the MEA and amino-silicone solvent (Case 1) baseline cases are provided in Exhibits 1-1 and 1-2 respectively.

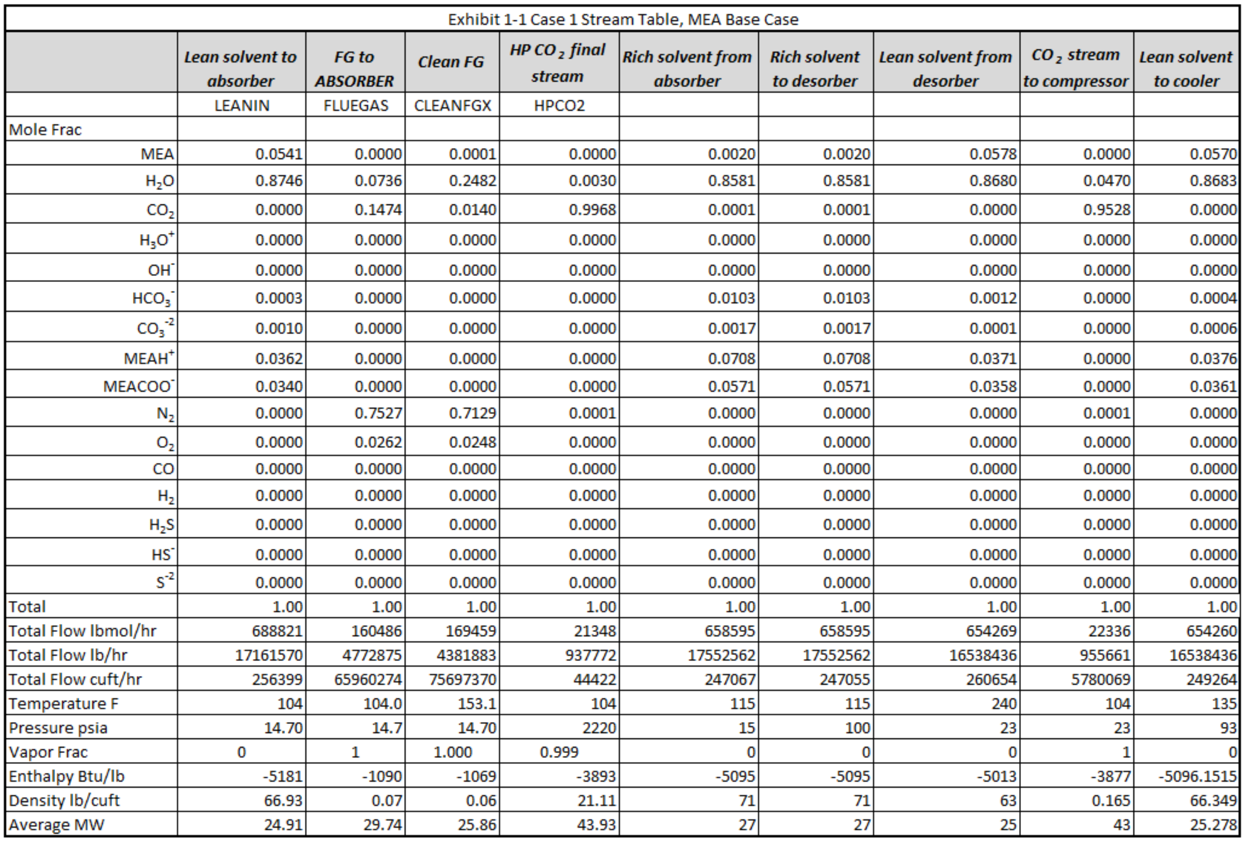


Exhibit 1-1 Case 1 Stream Table, MEA Base Case

	Lean solvent to absorber	*FG to ABSORBER*	*Clean FG*	*HP CO_2 final stream*	*Rich solvent from absorber*	*Rich solvent to desorber*	*Lean solvent from desorber*	*CO_2 stream to compressor*	*Lean solvent to cooler*
	LEANIN	FLUEGAS	CLEANFGX	HPCO2					
Mole Frac									
MEA	0.0541	0.0000	0.0001	0.0000	0.0020	0.0020	0.0578	0.0000	0.0570
H_2O	0.8746	0.0736	0.2482	0.0030	0.8581	0.8581	0.8680	0.0470	0.8683
CO_2	0.0000	0.1474	0.0140	0.9968	0.0001	0.0001	0.0000	0.9528	0.0000
H_3O^+	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
OH^-	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HCO_3^-	0.0003	0.0000	0.0000	0.0000	0.0103	0.0103	0.0012	0.0000	0.0004
CO_3^{-2}	0.0010	0.0000	0.0000	0.0000	0.0017	0.0017	0.0001	0.0000	0.0006
$MEAH^+$	0.0362	0.0000	0.0000	0.0000	0.0708	0.0708	0.0371	0.0000	0.0376
$MEACOO^-$	0.0340	0.0000	0.0000	0.0000	0.0571	0.0571	0.0358	0.0000	0.0361
N_2	0.0000	0.7527	0.7129	0.0001	0.0000	0.0000	0.0000	0.0001	0.0000
O_2	0.0000	0.0262	0.0248	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
CO	0.0000	0.0000	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	0.0000	0.0000
H_2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
HS^-	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
S^{-2}	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
Total	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Flow lbmol/hr	688821	160486	169459	21348	658595	658595	654269	22336	654260
Total Flow lb/hr	17161570	4772875	4381883	937772	17552562	17552562	16538436	955661	16538436
Total Flow cuft/hr	256399	65960274	75697370	44422	247067	247055	260654	5780069	249264
Temperature F	104	104.0	153.1	104	115	115	240	104	135
Pressure psia	14.70	14.7	14.70	2220	15	100	23	23	93
Vapor Frac	0	1	1.000	0.999	0	0	0	1	0
Enthalpy Btu/lb	-5181	-1090	-1069	-3893	-5095	-5095	-5013	-3877	-5096.1515
Density lb/cuft	66.93	0.07	0.06	21.11	71	71	63	0.165	66.349
Average MW	24.91	29.74	25.86	43.93	27	27	25	43	25.278

Exhibit 1-2 Stream Table, GAP-1/TEG Base Case

	Lean solvent into absorber	***FG to absorber***	***Clean FG***	***HP CO_2 final stream***	***Rich solvent from absorber***	***Rich Solvent to desorber***	***Lean solvent from desorber***	***Lean solvent to cooler***	***CO_2 to compressor***
Mole Frac									
H_2O	0.2407	0.0731	0.0659	0.0008	0.2583	0.2669	0.2350	0.2350	0.0170
CO_2	0.0082	0.1475	0.0165	0.9919	0.0005	0.0005	0.0084	0.0084	0.9759
N_2	0.0001	0.7532	0.8861	0.0072	0.0012	0.0012	0.0001	0.0001	0.0071
O_2	0.0000	0.0262	0.0309	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
GAP-S	0.2653	0.0000	0.0004	0.0000	0.0912	0.0906	0.2632	0.2632	0.0000
GAP-S CARB	0.0339	0.0000	0.0003	0.0000	0.2032	0.2007	0.0377	0.0377	0.0000
D4	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000
TEG	0.4519	0.0000	0.0000	0.0000	0.4456	0.4400	0.4556	0.4556	0.0000
Total	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.00
Total Flow lbmol/hr	130000	160395	136160	21065	131821	133536	128942	128942	21415
Total Flow lb/hr	22148650	4771236	3803405	924170	23116261	23179467	22124069	22124069	930560
Total Flow cuft/hr	302858	66001755	59958986	59567	309581	333217	339703	315827	2056175
Temperature F	104	104	143	124	158	260	285	179	104
Pressure psia	15	14.7	14.6959488	2215	15	63	63	63	63
Vapor Frac	0	1	1	1	0	0	0	0	1
Enthalpy Btu/lb	-2241.92	-1087.77	-343.17	-3817.30	-2334	-2288	-2153	-2206	-3834
Density lb/cuft	73.13	0.07	0.06	15.51	75	70	65	70	0.45
Average MW	170	30	28	44	175	174	172	172	43

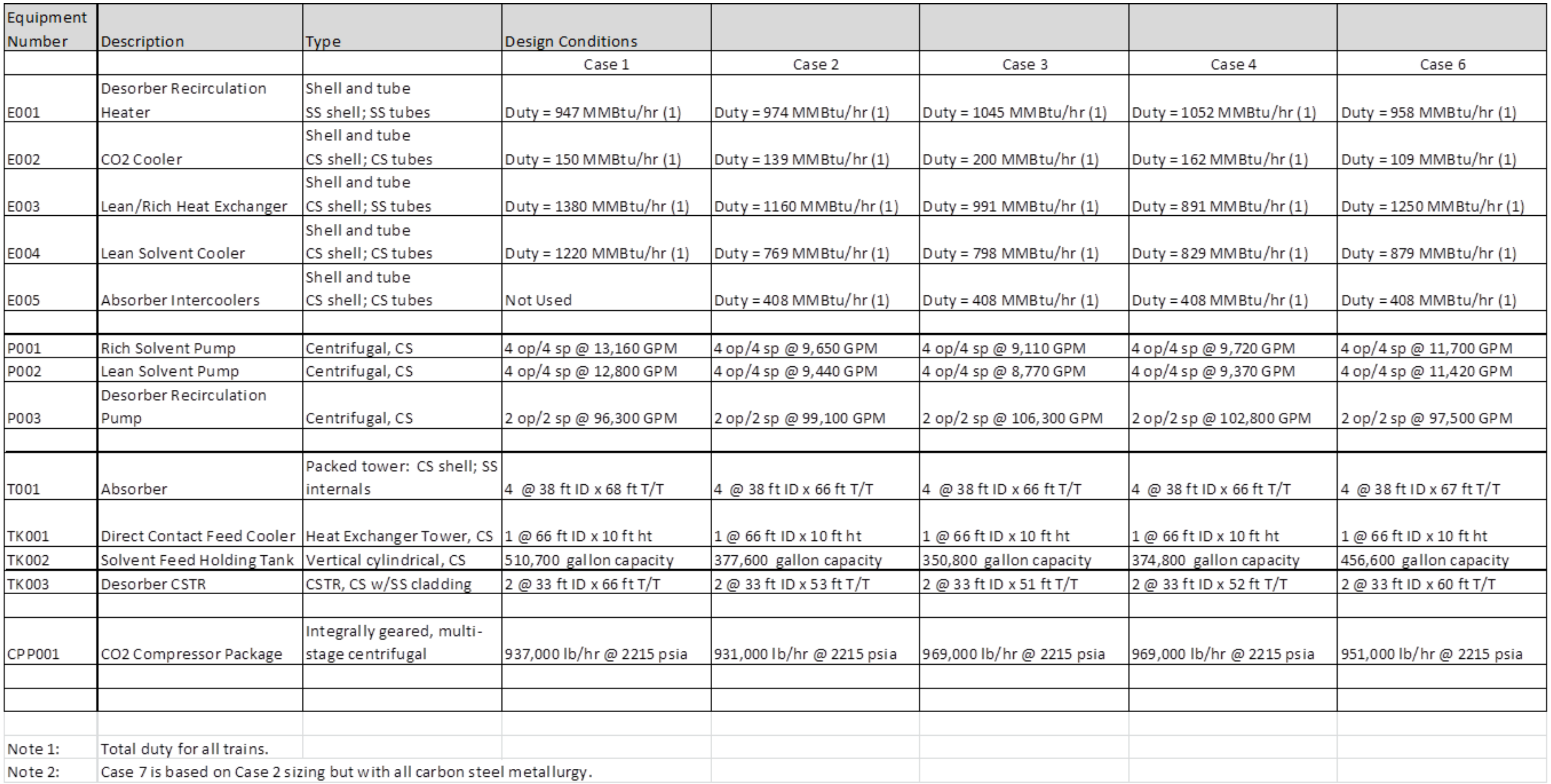
The MEA equipment list is summarized in the following table.

Table 7. MEA equipment list.

Equipment Number	***Description***	***Type***	***Design Conditions***	***Quanitity***
NA	Desorber recirculation heater	Not used	NA	NA
E002	CO_2 Cooler	Shell and tube; CS Shell; CS Tubes	Heat duty = 424 MMBTU/hr (1)	4
E003	Rich/Lean heat Exchanger	Shell and tube; CS shell; SS Tubes	Heat duty = 1,413 MMBTU/hr	1
E004	Lean Solvent cooler	Shell and tube; CS Shell; CS Tubes	Heat duty = 691 MMBTU/hr	1
NA	Absorber intercoolers	Not used	NA	NA
P001	Rich Solvent Pump	Centrifugal, CS	9,140 gpm at 85 psi pressure change	4
P002	Lean Solvent Pump	Centrifugal, CS	9,593gpm at 69.7 psi pressure change	4
	Desorber Recirculation pump	Not used	NA	NA
T001	Absorber	Packed Tower	39ft ID x 65.6 ft packed height	4
TK001	Direct contact Feed cooler	Heat Exchanger Tower, CS	66 ft ID x 10 ft height	1
TK002	Solvent feed holding tank	Vertical Cylindrical, CS	319675 gallon capacity	1
T002	Desorber	Packed Tower	25ft ID x 65.6 ft packed height	4
CPP001	CO_2 compression package	Integrally geared, multistage centrifugal	937,000 lb/hr at 2215 psia	1 train

The GAP-1/TEG equipment table is summarized below.

Table 8. GAP-1/TEG equipment list.



Equipment Number	Description	Type	Design Conditions				
			Case 1	Case 2	Case 3	Case 4	Case 6
E001	Desorber Recirculation Heater	Shell and tube SS shell; SS tubes	Duty = 947 MMBtu/hr (1)	Duty = 974 MMBtu/hr (1)	Duty = 1045 MMBtu/hr (1)	Duty = 1052 MMBtu/hr (1)	Duty = 958 MMBtu/hr (1)
E002	CO2 Cooler	Shell and tube CS shell; CS tubes	Duty = 150 MMBtu/hr (1)	Duty = 139 MMBtu/hr (1)	Duty = 200 MMBtu/hr (1)	Duty = 162 MMBtu/hr (1)	Duty = 109 MMBtu/hr (1)
E003	Lean/Rich Heat Exchanger	Shell and tube CS shell; SS tubes	Duty = 1380 MMBtu/hr (1)	Duty = 1160 MMBtu/hr (1)	Duty = 991 MMBtu/hr (1)	Duty = 891 MMBtu/hr (1)	Duty = 1250 MMBtu/hr (1)
E004	Lean Solvent Cooler	Shell and tube CS shell; CS tubes	Duty = 1220 MMBtu/hr (1)	Duty = 769 MMBtu/hr (1)	Duty = 798 MMBtu/hr (1)	Duty = 829 MMBtu/hr (1)	Duty = 879 MMBtu/hr (1)
E005	Absorber Intercoolers	Shell and tube CS shell; CS tubes	Not Used	Duty = 408 MMBtu/hr (1)	Duty = 408 MMBtu/hr (1)	Duty = 408 MMBtu/hr (1)	Duty = 408 MMBtu/hr (1)
P001	Rich Solvent Pump	Centrifugal, CS	4 op/4 sp @ 13,160 GPM	4 op/4 sp @ 9,650 GPM	4 op/4 sp @ 9,110 GPM	4 op/4 sp @ 9,720 GPM	4 op/4 sp @ 11,700 GPM
P002	Lean Solvent Pump	Centrifugal, CS	4 op/4 sp @ 12,800 GPM	4 op/4 sp @ 9,440 GPM	4 op/4 sp @ 8,770 GPM	4 op/4 sp @ 9,370 GPM	4 op/4 sp @ 11,420 GPM
P003	Desorber Recirculation Pump	Centrifugal, CS	2 op/2 sp @ 96,300 GPM	2 op/2 sp @ 99,100 GPM	2 op/2 sp @ 106,300 GPM	2 op/2 sp @ 102,800 GPM	2 op/2 sp @ 97,500 GPM
T001	Absorber	Packed tower: CS shell; SS internals	4 @ 38 ft ID x 68 ft T/T	4 @ 38 ft ID x 66 ft T/T	4 @ 38 ft ID x 66 ft T/T	4 @ 38 ft ID x 66 ft T/T	4 @ 38 ft ID x 67 ft T/T
TK001	Direct Contact Feed Cooler	Heat Exchanger Tower, CS	1 @ 66 ft ID x 10 ft ht	1 @ 66 ft ID x 10 ft ht	1 @ 66 ft ID x 10 ft ht	1 @ 66 ft ID x 10 ft ht	1 @ 66 ft ID x 10 ft ht
TK002	Solvent Feed Holding Tank	Vertical cylindrical, CS	510,700 gallon capacity	377,600 gallon capacity	350,800 gallon capacity	374,800 gallon capacity	456,600 gallon capacity
TK003	Desorber CSTR	CSTR, CS w/SS cladding	2 @ 33 ft ID x 66 ft T/T	2 @ 33 ft ID x 53 ft T/T	2 @ 33 ft ID x 51 ft T/T	2 @ 33 ft ID x 52 ft T/T	2 @ 33 ft ID x 60 ft T/T
CPP001	CO2 Compressor Package	Integrally geared, multi-stage centrifugal	937,000 lb/hr @ 2215 psia	931,000 lb/hr @ 2215 psia	969,000 lb/hr @ 2215 psia	969,000 lb/hr @ 2215 psia	951,000 lb/hr @ 2215 psia

Note 1: Total duty for all trains.
Note 2: Case 7 is based on Case 2 sizing but with all carbon steel metallurgy.

Cost Estimates

In Phase 1 of the project in 2012, cost estimates were conducted using the Aspen^{TM} Cost Estimator using the results from the revised Aspen Plus^{TM} models. The same Aspen software was used for the current study with some minor refinements made to the 2012 model. The 2012 model had used the default plot plan which was much too small. For 2013, the plot plan area was estimated using sized equipment dimensions when available. The final plot plan area was 72,600 ft^2.

In the 2012 report, the importance of the lean/rich heat exchanger heat transfer coefficient was pointed out due to the significant of the lean/rich solvent heat exchanger area on the overall costs. Refer to the 2012 report for the detailed discussion. The same heat-transfer coefficients used in the 2012 report were used for this phase, namely, 75 $Btu/hr/ft^2/°F$ for the GAP-1/TEG amino-silicone solvent and 120 $Btu/hr/ft^2/°F$ for the MEA solvent.

For the MEA model, a six stage compressor using the same inter-stage pressures as the one described in Section 4.1.7 of the DOE/NETL-2007/1281 report was used. Inter-stage coolers and knockout drums for the compressor were also sized. Carbon steel metallurgy was assumed.

For GAP-1/TEG, the absorber was still sized as a packed tower but the desorber is now modeled and sized as a CSTR reactor. For MEA, the DOE/NETL-2007/1281 report had used packed towers so packed towers were used for both of these vessels in the MEA cost estimate.

For MEA, the rich solvent is corrosive so the absorber towers were assumed to use carbon steel shells with stainless steel packing. Also, stainless steel tubes were used for the lean/rich solvent heat exchanger and the steam reboiler in the desorber column.

In the first half of 2013, bench-scale experimental work on the amino-silicone system was conducted. Consequently, more accurate amino-silicone solvent circulation rates were used to re-size the equipment. For example, compared to the conditions for Case 1 (used in 2012 report), the solvent circulation mass rate was increased by a factor of 1.5.

For the amino-silicone (GAP-1/TEG) solvent, six cases are considered (Cases 1-4, 6 and 7). For Cases 1-4 and 6, the metallurgy for the carbon separation unit was assumed to be carbon steel with the exception of heat exchanger tubes in the hot, rich solvent service and column internals (packing), which were made of stainless steel. In Case 7, all parts are assumed to be carbon steel to check the sensitivity of the cost of the metallurgy requirement. It is expected that the amino-silicone system is less corrosive than the MEA system and further studies will need to be done to evaluate if all components can be made from carbon steel.

The features of the six cases are shown in Table 9.

Table 9. Amino-silicone case features.

	Case 1	Case 2	Case 3	Case 4	Case 6	Case 7
Absorber Intercooler	No	Yes	Yes	Yes	Yes	Yes
Desorber Pressure, psia	63	63	20	20	63	63
Desorber Temp, deg C	140	140	140	130	130	140
Material	CS/SS	CS/SS	CS/SS	CS/SS	CS/SS	CS

Amino-Silicone (GAP-1/TEG) System

The capital cost for the MEA unit for Case 10 in the DOE/NETL-2007/1281 report is $484 Million and this is based on flue gas flow of 7,578,830 lbs/hr. In this contract, DOE specified that the flue gas flow for this study is 5,118,399 lbs/hr. The correction factor for capital cost based on capacity differences between the DOE report and this contract is:

CAPEX Capacity Correction Factor= (5,118,399 lbs per hour/ 7,578,830 lbs per hour)$^{0.6}$ = 0.79

Therefore, the capacity adjusted capital cost for the MEA unit based on DOE NETL Report is:

= 0.79 x $484.5 Million ~ $383 Million

GE also did an estimate of an MEA unit as an independent check and the resulting project cost was in relatively good agreement.

The relative capital costs of the various amino-silicone solvent cases versus the conventional MEA case (based on modeling performed by GE in this contract, using commercially available Aspen^{TM} cost estimation software) are given in the Table 10.

Table 10. Relative CAPEX case summary.

	Case 1	Case 2	Case 3	Case 4	Case 6	Case 7
Relative Cost to MEA	0.93	0.86	0.87	0.82	0.89	0.61

Case 1 is operated at the same absorber/desorber conditions as reported in the 2012 report. Overall project cost rose by approximately 10%. The chief driver for the cost increase is the increased solvent circulation rate (about 50% higher) which results in higher heat exchange costs. Another cost adder was the switch to a CSTR type reactor system. On the other hand, costs were mitigated downward by optimizing the absorber system (reduced packed bed and overall absorber height).

Case 2 clearly demonstrates the favorable impact of absorber intercooling. Absorber intercooling is kept for the rest of the cases.

Cases 3 and 4 operate at a lower desorber pressure of 20 psia. This results in a reduced solvent circulation requirement which is off-set by the increasing product CO_2 compression cost (another compression stage is needed).

Case 6 is similar to Case 2 with the exception of a lower desorber temperature. Higher CO_2 concentration in the lean solvent to the absorber (reduced solvent capacity) results in a higher solvent circulation rate. The overall impact is a CAPEX increase.

Case 7 assumes all carbon steel metallurgy to check the sensitivity of the cost of metallurgy requirement. The cost of amino-silicone solvent Case 2 was repeated assuming a lower metallurgy for the tower internals and heat exchanger tubes (substituting carbon steel for stainless steel). It is expected that the amino-silicone system is less corrosive than the MEA system and further studies will need to be done to evaluate if all components can be made from carbon steel. If only carbon steel is required, the overall unit costs for the amino-silicone system would fall to ~70% of the amino-solvent case and to ~60% of the MEA case.

Power Plant Efficiency and Energy Penalty

The system utilities for the MEA system are summarized in the following table:

POWER SUMMARY	
AUXILIARY LOAD SUMMARY, kWe	
Feed Gas Blower	9,254
CO_2 Separation Auxiliaries	2,896
CO_2 Compression	39,712
Cooling Water Fans/Pumps	7,242
TOTAL AUXILIARIES, kWe	59,104
COOLING WATER, ton/hr	19,051
STEAM, ton/hr	688.3

The system utilities for a typical GAP-1/TEG system are summarized in the following table:

POWER SUMMARY	
AUXILIARY LOAD SUMMARY, kWe	
Feed Gas Blower	9,254
CO_2 Separation Auxiliaries	2,087
CO_2 Compression	35,208
Cooling Water Fans/Pumps	14,332
TOTAL AUXILIARIES, kWe	60,881
COOLING WATER, ton/hr	37,704
STEAM, ton/hr	594.1

The water falls for power plant net efficiency due to CCS are shown in Figure 28 and Figure 29 at steam temperatures of 743 °F and 400 °F, respectively. The water falls for power plant energy penalty due to CCS are shown in Figure 30 and Figure 31 at steam temperatures of 743 °F and

400 °F, respectively. The 1st column is for the MEA solvent. The remaining columns are for amino-silicone solvent cases. The amino-silicone cases have significantly lower energy penalties when compared with MEA.

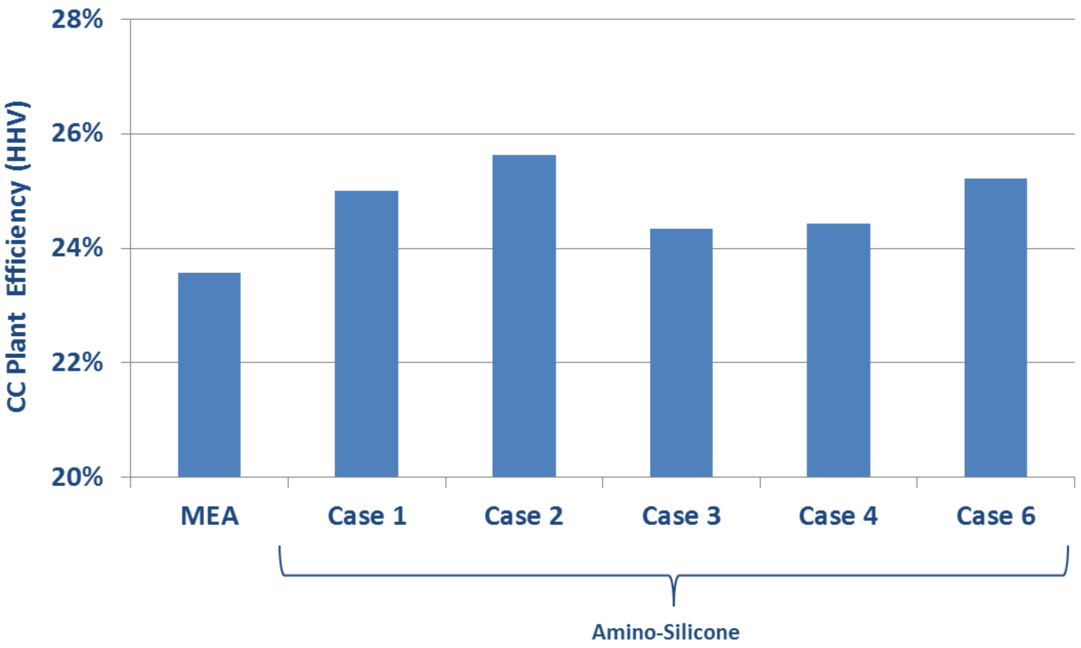


Figure 28. Plant Efficiency Based on HHV at steam temperatures of 743 °F.

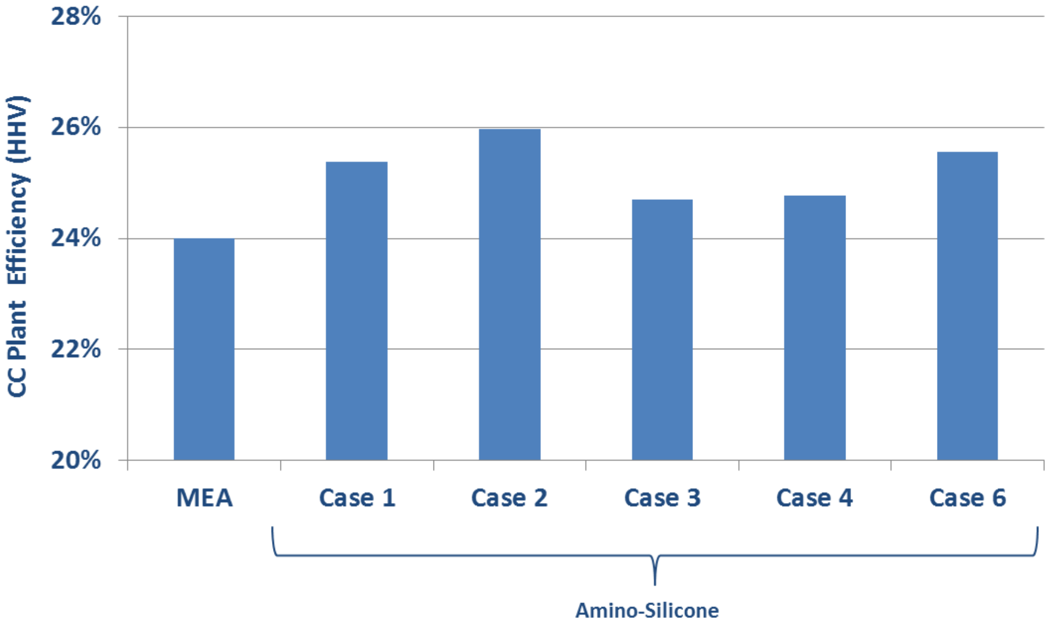


Figure 29. Plant Efficiency Based on HHV at steam temperatures of 400 °F.

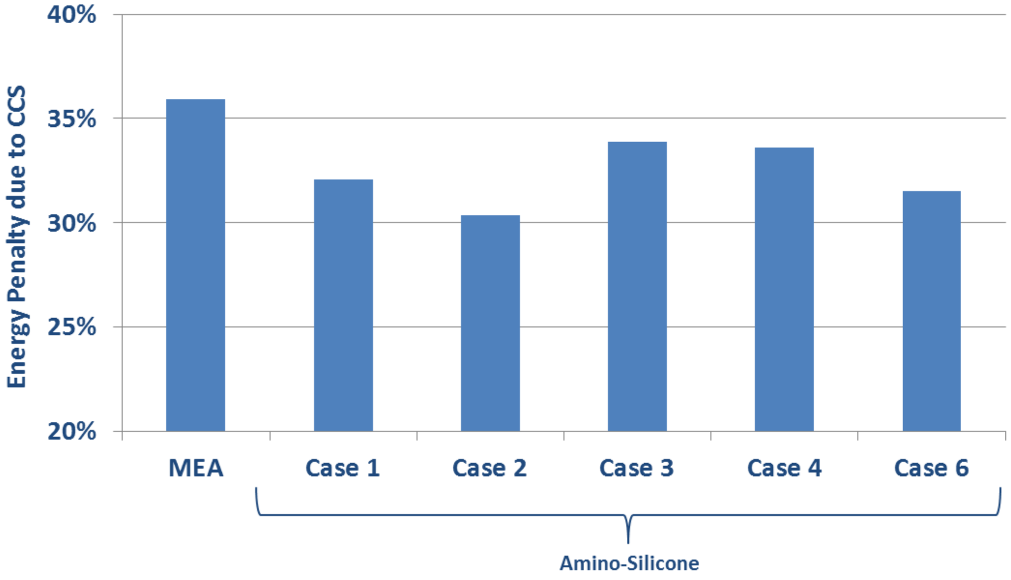


Figure 30. Energy penalty due to CCS at steam temperatures of 743 °F

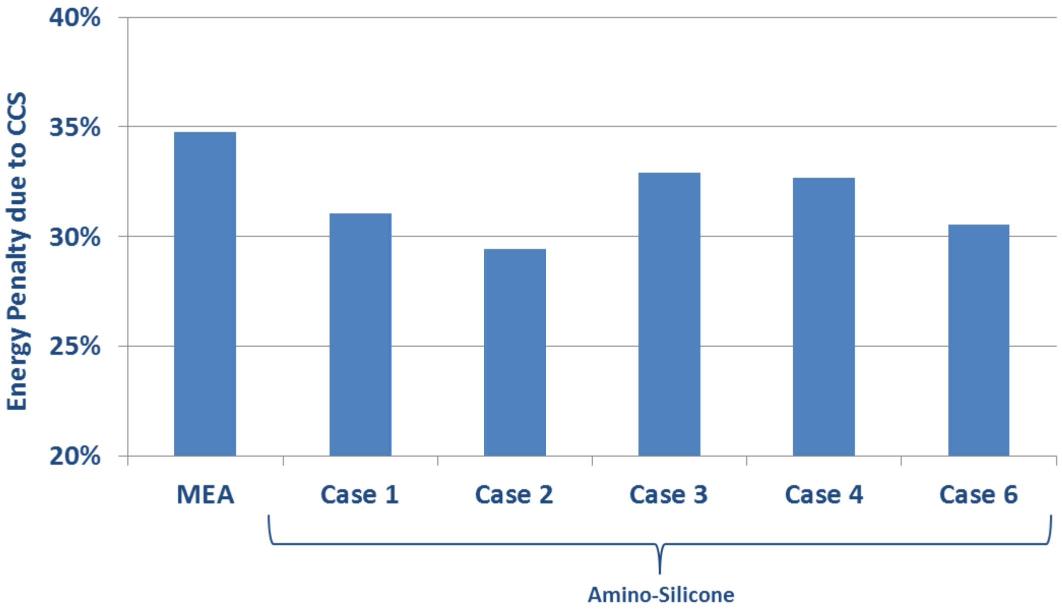


Figure 31. Energy penalty due to CCS at steam temperatures of 400 °F.

Cost of Electricity

<u>Economic Analysis Assumptions:</u>

As per the cooperative agreement the following economic assumptions were used:
1) Levelized-Cost of Electricity without CO_2 capture: 64 mills/kWh
2) Levelized-Cost of Steam without CO_2 capture: $5.83 / 1,000 lbs
3) Levelized-Cost of Cooling Water: $0.12 / 1,000 gal
4) Levelized-Cost of Process Make-Up Water: $0.07 / 1,000 lb
5) Levelized-Cost of Wastewater Treatment: $0.21 / lb
6) Levelized-Cost of Solid-Waste Disposal: $17.87 / ton
7) Levelized-Cost of Toxic-Waste Disposal: $89.36 / ton
8) Levelized-Cost of CO_2 Transport, Storage & Monitoring: $4.05 / ton CO_2
9) Plant On-Stream Factor: 310.25 days/yr
10) Retrofit Factor: 1.0
11) Plant Location: generic plant site, U.S. Midwest
12) Dollar-Year Reporting Basis: 2007
13) Total Fixed O&M Levelized-Costs $995 / calendar day
14) Levelized Maintenance-Material Costs 2% (as percentage of initial equipment and materials costs)

15) Capital Charge Factor: 17.5%/yr (based on 20-year levelized cost of electricity, LCOE)

The solvent usage per year was calculated assuming that one mole of SO_2 degrades one mole of solvent. This resulted in a GAP-1 loss of 122 lbs/hr.

The water fall chart for increase in cost of electricity (COE) over a non-capture case is shown in Figure 32 for a steam temperature of 743 °F, using the following equation which is specified in the contract and is marked “simple” in the chart.

$$\{\text{cost of electric power in mills/kWh}\} = 0.3073 \times \{\text{total power loss in MWe}\} + 64.00$$

The water fall chart for increase in COE using the “simple” calculation is shown in Figure 33 for a steam temperature of 400 °F. The best amino-silicone case has significantly lower COE when compared with MEA.

As outlined earlier, a range for the solvent cost was estimated by SiVance. Using the DOE assumptions listed above, the increase in COE, at steam temperatures of 743 °F and the higher solvent cost, over a plant without CO_2 capture is shown in Figure 34. This figure is marked “detailed” since more detailed calculations were performed. The increase in COE, at steam temperatures of 743 °F and the lower solvent cost, over a plant without CO_2 capture at the low solvent cost is shown in Figure 35. The increase in COE at steam temperatures of 400 °F and different solvent costs is shown in Figure 36 and Figure 37.

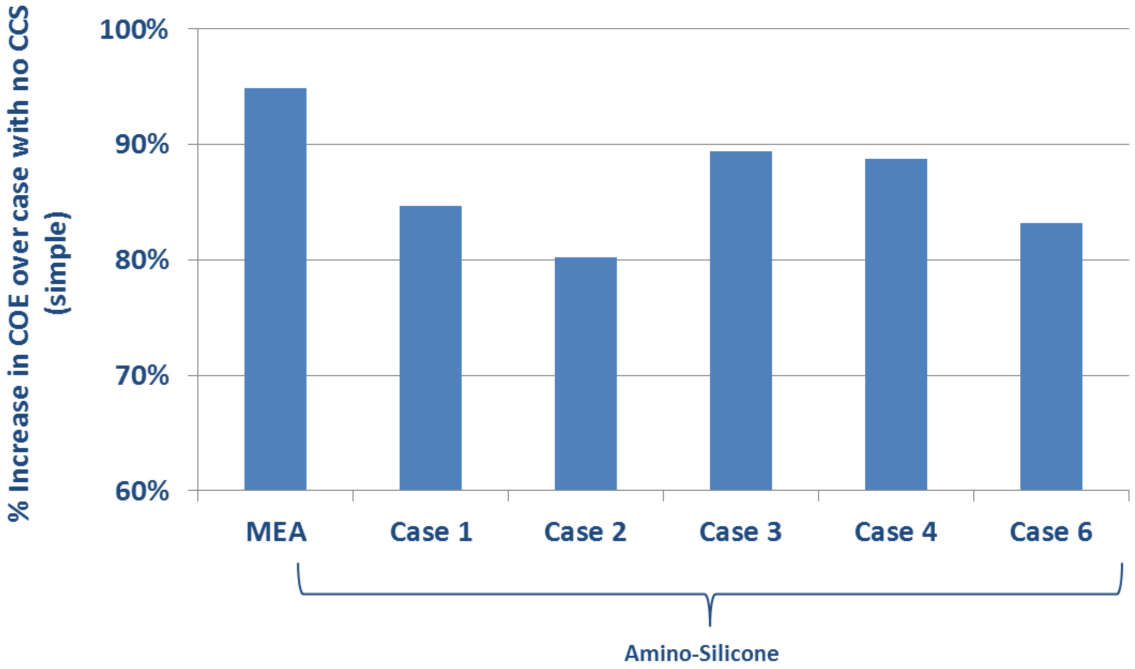


Figure 32. Increase in COE using simplified calculation as set in the contract at a steam temperature of 743 °F.

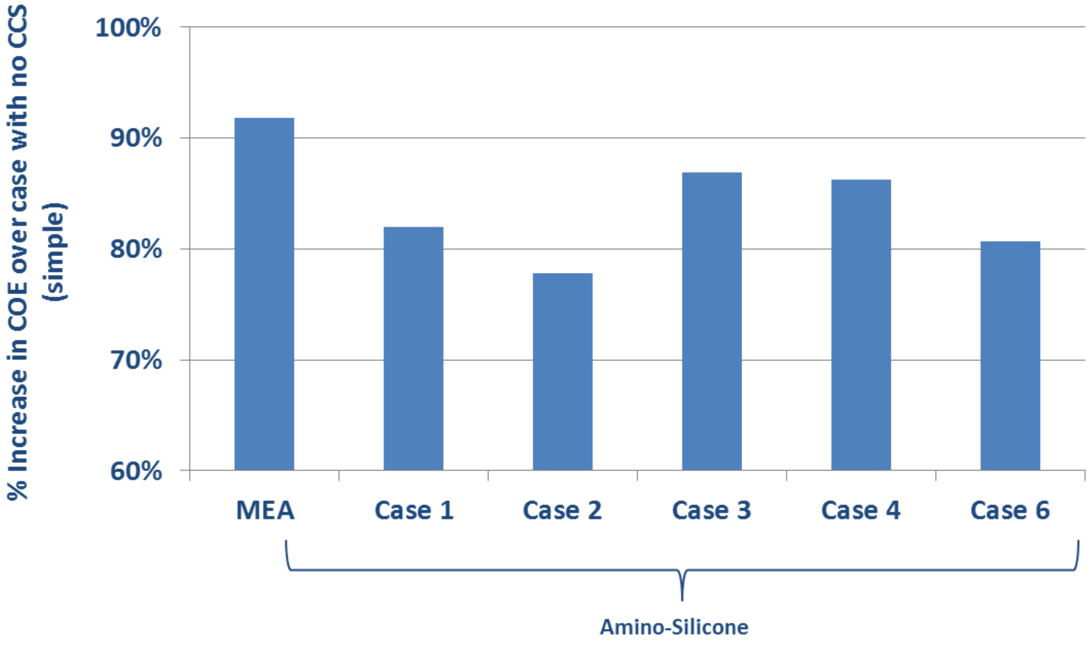


Figure 33. Increase in COE using simplified calculation as set in the contract at a steam temperature of 400 °F.

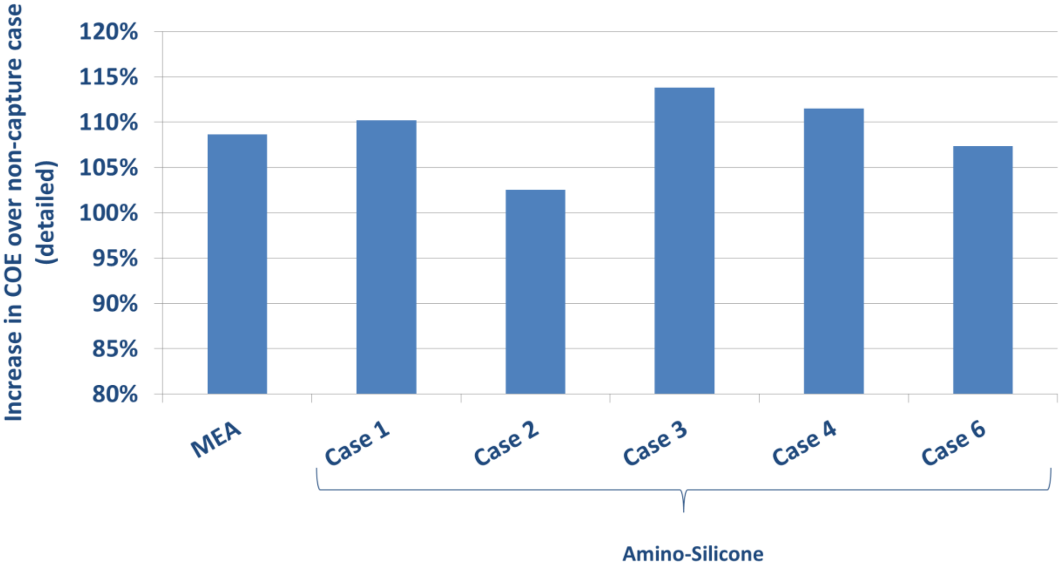


Figure 34. Increase in COE as calculated from detailed analysis using energy penalty and Capex estimates at a steam temperature of 743 °F and the high solvent cost.

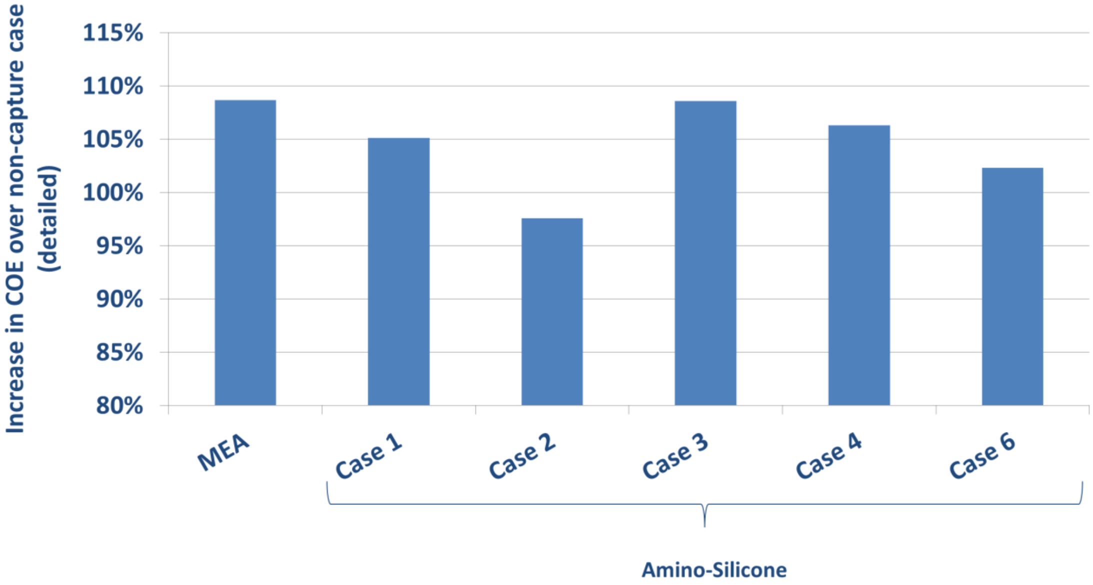


Figure 35. Increase in COE as calculated from detailed analysis using energy penalty and Capex estimates at a steam temperature of 743 °F and the low solvent cost.

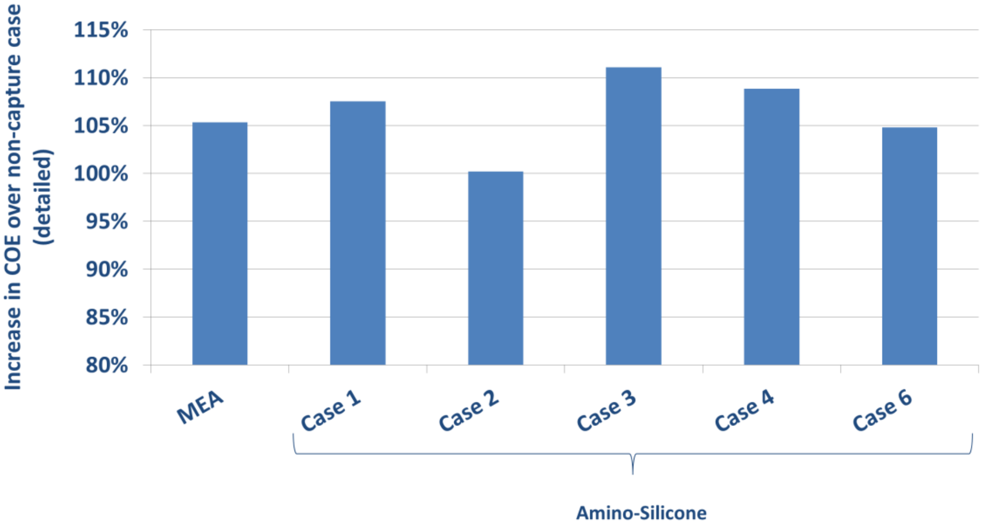


Figure 36. Increase in COE as calculated from detailed analysis using energy penalty and Capex estimates at a steam temperature of 400 °F and the high solvent cost.

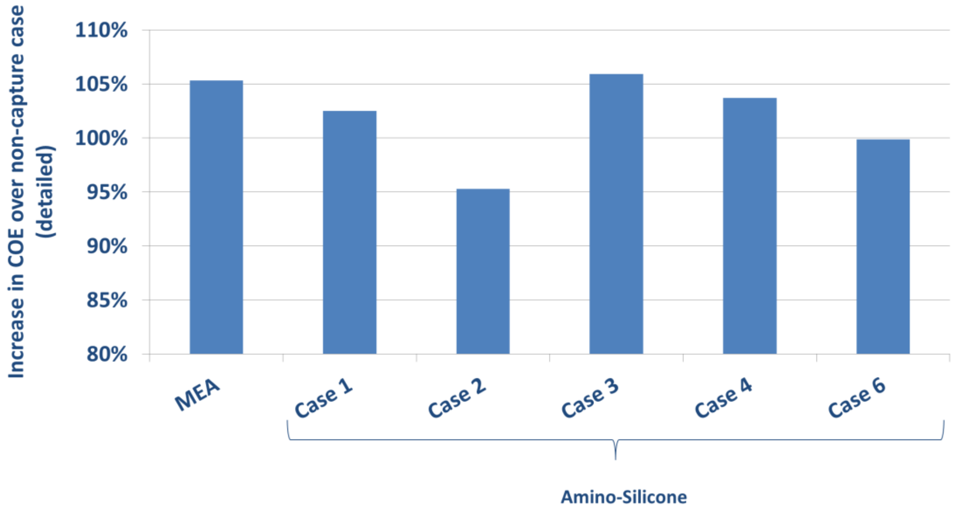


Figure 37. Increase in COE as calculated from detailed analysis using energy penalty and Capex estimates at a steam temperature of 400 °F and the low solvent cost.

As seen in the economic analysis above, Case 2 consistently outperforms the MEA case.

The cost of steam is shown in the following tables using the following equation, specified by DOE.

Cost of steam in \$/1000 lb = 0.028 * {Total power loss in MWe} + 5.83

The cost of CO_2 is shown in the following tables using 3 different methodologies, specified by DOE.

1. Cost of CO_2 method specified in the cooperative agreement, which is
   Total cost of capturing CO_2 =
   {cost of all materials and utilities consumed in the CO_2 system}
   + {cost of treating or disposing of any effluent streams from the system, including transport, storage and monitoring of CO_2}
   + {cost of maintenance and materials}
   + {fixed O&M Costs}
   + {capital charge for CO_2 capture system}
   The cost per ton of CO_2 is obtained by dividing this cost by the tons of CO_2 captured by the process.

2. Removal cost of CO_2, specified in bituminous baseline report

$$Removal\ Cost = \frac{\{LCOE_{with\ removal} - LCOE_{w/o\ removal}\}\$/MWh}{\{CO_2\ removed\}tons/MWh}$$

3. Avoided cost of CO_2, specified in bituminous baseline report

$$Avoided\ Cost = \frac{\{LCOE_{with\ removal} - LCOE_{w/o\ removal}\}\$/MWh}{\{Emissions_{w/o\ removal} - Emissions_{with\ removal}\}tons/MWh}$$

A summary of key parameters specified by DOE are shown in the following table for GAP-1/TEG Case 2 with the higher solvent cost.

	MEA	GAP1-TEG
Cost of Steam - $/1000 lbs	$ 11.36	$ 10.51
Cost of CO2 - $/ton	$ 25.42	$ 30.95
Removal cost for CO2 - $/ton	$ 52.53	$ 53.87
Avoided Cost for CO2 - $/ton	$ 87.31	$ 69.55
% decrease in PC Plant Efficiency	13.2%	11.2%

A summary of key parameters specified by DOE are shown in the following table for GAP-1/TEG Case 2 with the lower solvent cost.

	MEA	GAP1-TEG
Cost of Steam - $/1000 lbs	$ 11.36	$ 10.51
Cost of CO2 - $/ton	$ 25.42	$ 28.34
Removal cost for CO2 - $/ton	$ 52.53	$ 51.26
Avoided Cost for CO2 - $/ton	$ 87.31	$ 66.18
% decrease in PC Plant Efficiency	13.2%	11.2%

Conclusions:

System and economic analysis for a carbon capture unit which uses an amino-silicone solvent for CO_2 capture and sequestration (CCS) in a pulverized coal (PC) boiler demonstrates that the amino-silicone solvent has significant advantages relative to an MEA-based system. The CCS energy penalty for MEA is 35.9% and the energy penalty for amino-silicone solvent is 30.4% using a steam temperature of 395 °C (743 °F). If the steam temperature is lowered to 204 °C (400 °F), the energy penalty for the amino-silicone solvent is reduced to 29%. The increase in

COE over the non-capture case for MEA is ~109% and increase in COE for amino-silicone solvent is ~98 to 103% depending on the solvent cost at a steam temperature of 395 °C (743 °F). If the steam temperature is lowered to 204 °C (400 °F), the increase in COE for the amino-silicone solvent is reduced to ~95-100%.