Bench-Scale Silicone Process for Low-Cost CO_2 Capture

Preliminary Technical and Economic Feasibility Report

Topical Report

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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. The report also shows results for a CCS unit based on a conventional approach using mono-ethanol amine (MEA). Models were developed for both processes and used to calculate mass and energy balances. Capital costs and energy penalty were calculated for both systems, as well as the increase in cost of electricity. The amino-silicone solvent based system demonstrates significant advantages compared to the MEA system.

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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 unit 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 cooperative agreement (DE-FE0007502).

At a steam temperature of 395 °C (743 °F), the CCS energy penalty for amino-silicone solvent is only 23.7% which compares to a 35.6% energy penalty for MEA. At a lower steam temperature of 171 °C (340 °F), the energy penalty for the amino-silicone solvent is 23.0%.

At a steam temperature of 395 °C (743 °F), the increase in COE for amino-silicone solvent is 86% which compares to an ~110% COE cost increase for MEA. At a lower steam temperature of 171 °C (340 °F), the increase in COE for the amino-silicone solvent is 85%. In summary, the amino-silicone solvent has a significant advantage over conventional systems using MEA.

Completion of Task 2.2: Preliminary 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.

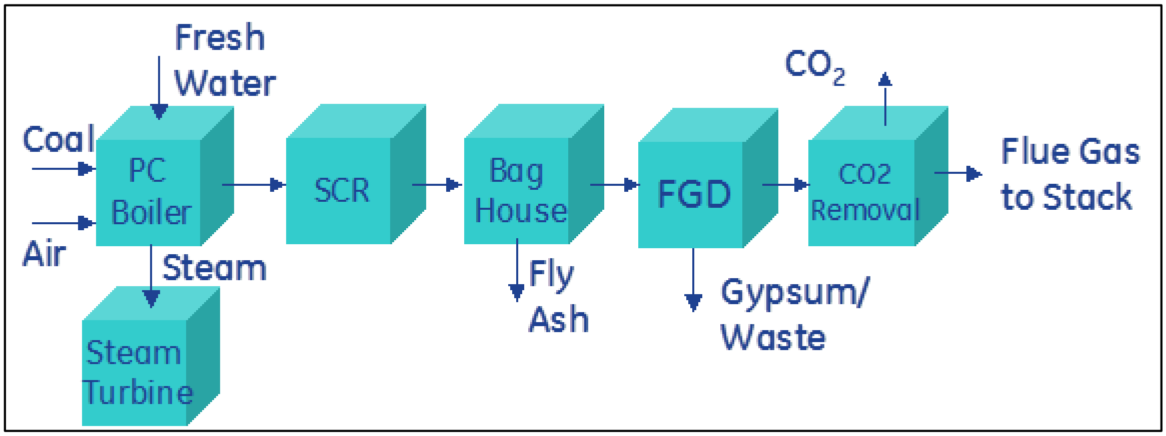

Figure 1: System block diagram.

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.

For the amino-silicone solvent baseline case, the desorber operates at 143 °C (290 °F) and 4.3 atm (63 psia). For the sensitivity studies, the desorber conditions were varied from 120 to 180 °C (248 to 356 °F) and from atmospheric to 5.1 atm (75 psia) and these results are presented in subsequent sections. For the MEA Baseline Case, the desorber 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.

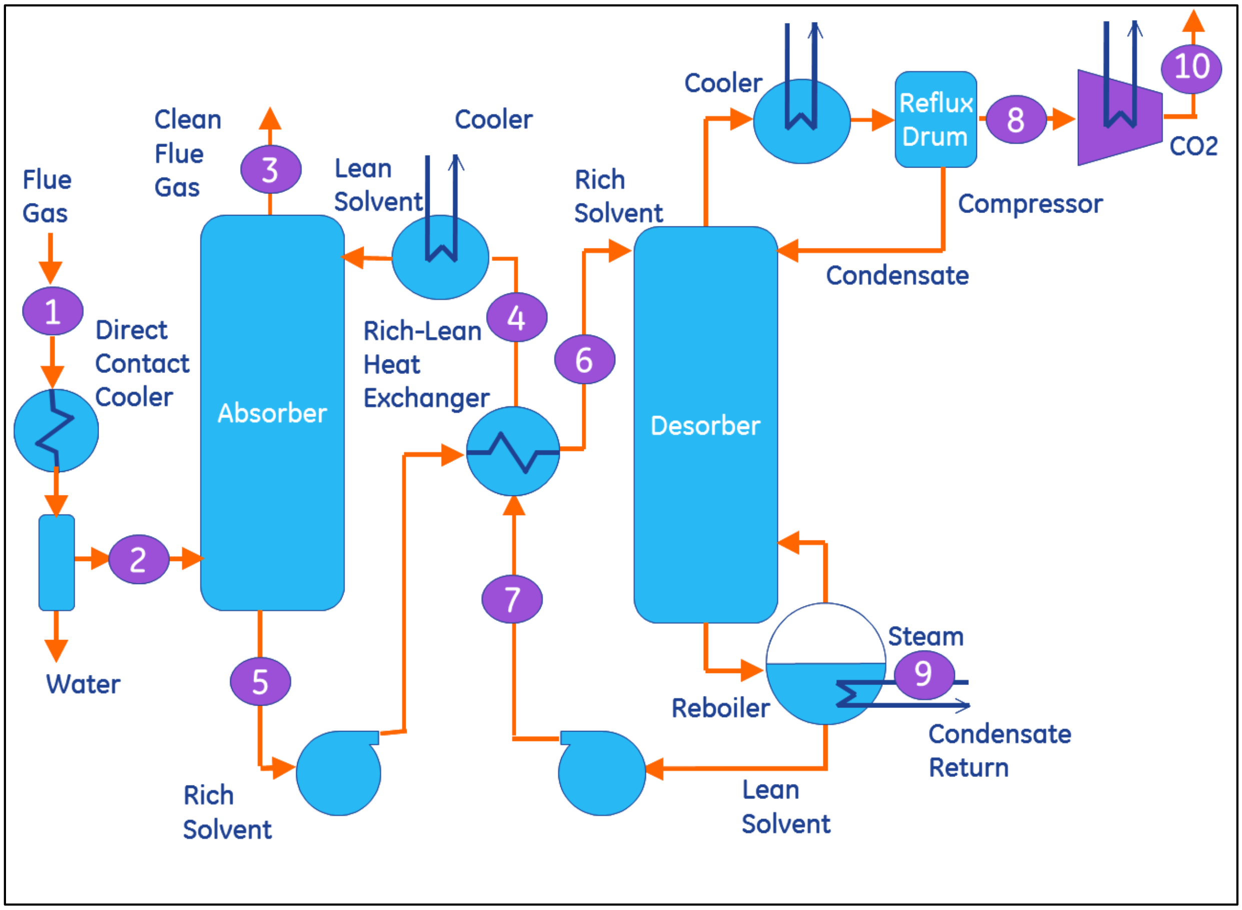

Figure 2: CO_2 separation sub-system.

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.

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 model accounts 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 column. The model also accounts 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 is obtained from lab-scale experiments for the GAP-1/TEG system. The lab-scale isotherm data indicate that sorbent net loading of 4.9% is achievable with GAP-1/TEG. Under the amino-silicone solvent baseline case conditions, a sorbent net loading of 4.1% (by weight) is achieved.

A detailed MEA Aspen model that was built previously under another program was used to compare the results for this study. The GE MEA model has a net loading of 3.7 wt%. Also, an Aspen Plus model for carbon capture using 60/40 wt% GAP-1/TEG system was built as shown in Figure 3.

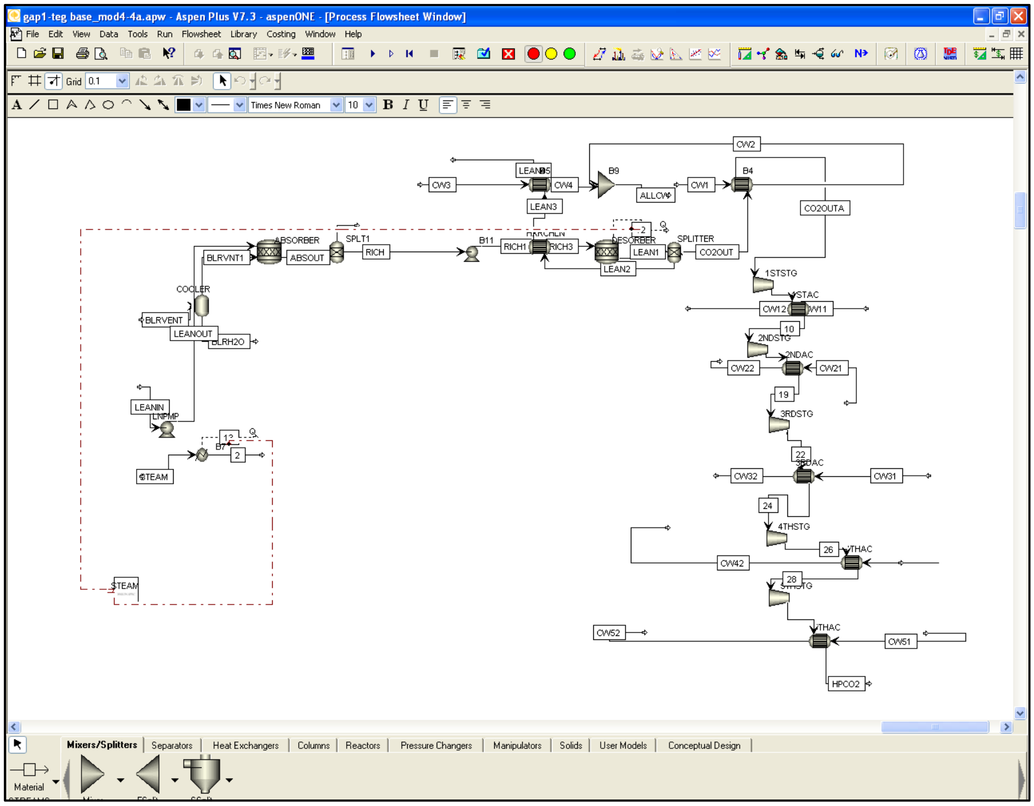

Figure 3: Aspen Plus model for carbon capture using GAP-1/TEG 60/40 wt%.

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 baseline case are provided in Exhibits 1-1 and 1-2 respectively.

The GAP-1/TEG sorbent utilizes less energy than the MEA sorbent due to:

- Low water in the sorbent mixture
- Low specific heat of the sorbent

The effect of these parameters is described in more detail below.

Low H_2O

The model accounts for absorption of water in the flue gas by the MEA sorbent and the vaporization of the water in the desorber column. The baseline MEA sorbent concentrations are limited to 20-30% and the remaining is water due to viscosity and corrosion issues. The water in the sorbent necessitates significant amount of energy due to sensible heat as well as vaporization of the water.

Low Specific Heat

The specific heat of GAP-1/TEG is 0.58 Btu/lb-F while the specific heat of MEA is 0.89 Btu/lb-F. The lower specific heat for GAP-1/TEG improves the energy efficiency.

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

	1	2	3	4	5	6	7	8	9	10
		FLUE GAS	CLEAN	LEAN	RICH	RICH	LEAN SOLV	CO2 TO	STEAM	HI PRESS
From	FEED	TO	FLUE GAS	SOLVENT	SOLVENT	SOLVENT	FROM	COMPR	TO	CO2
To		ABSORBER		TO COOLER	FRM ABS	TO DESORB	DESORBER		REBOILER	
Component Mole Fraction										
H2O	0.1725	0.0736	0.1962	0.8714	0.8679	0.8653	0.8714	0.0470	1.0000	0.0470
CO2	0.1317	0.1474	0.0351	0.0000	0.0000	0.0016	0.0001	0.9525	0	0.9525
MEA	0.0000	0.0000	0.0002	0.0601	0.0092	0.0139	0.0605	0.0000	0	0.0000
N2	0.6644	0.7438	0.7338	0.0000	0.0000	0.0000	0.0000	0.0001	0	0.0001
O2	0.0234	0.0262	0.0258	0.0000	0.0000	0.0000	0.0000	0.0000	0	0.0000
AR	0.0080	0.0090	0.0088	0.0000	0.0000	0.0000	0.0000	0.0004	0	0.0004
MEAH+	0	0	0	0.0344	0.0617	0.0597	0.0341	0	0	0
MEACOO-	0	0	0	0.0332	0.0566	0.0538	0.0331	0	0	0
HCO3-	0	0	0	0.0005	0.0040	0.0056	0.0007	0	0	0
CO3-2	0	0	0	0.0004	0.0005	0.0001	0.0001	0	0	0
H3O+	0	0	0	0.0000	0.0000	0.0000	0.0000	0	0	0
OH-	0	0	0	0.0000	0.0000	0.0000	0.0000	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Mole Flow, LBMOLE/HR	1.79E+05	1.60E+05	1.62E+05	7.78E+05	7.80E+05	7.81E+05	7.79E+05	2.23E+04	8.46E+04	2.23E+04
Mass Flow, LB/HR	5.12E+06	4.77E+06	4.35E+06	1.95E+07	2.04E+07	2.04E+07	1.95E+07	9.53E+05	1.52E+06	9.53E+05
Volume Flow, CUFT/HR	7.76E+07	6.57E+07	7.23E+07	2.96E+05	2.95E+05	3.08E+05	3.09E+05	5.76E+06	6.44E+06	2.29E+04
Temperature, Deg F	135.0	104.0	151.4	142.5	122.2	217.5	241.0	104.0	743.3	124.0
Pressure, PSIA	14.70	14.70	14.70	83.00	14.98	90.00	24.90	23.21	167.00	2215.00
Vapor Fraction	1.0	1.0	1.0	0.0	0.0	0.0	0.0	1.0	1.0	1.0
Mass Enthalpy, BTU/LB	-1392.1	-1085.7	-962.6	-5120.1	-5120.7	-5040.1	-5036.1	-3875.5	-5466.6	-3978.2
Mass Density, LB/CUFT	0.066	0.073	0.060	65.808	69.220	66.398	63.103	0.165	0.237	41.582
Average Molecular Weight	28.58	29.85	26.83	25.01	26.20	26.16	25.01	42.79	18.02	42.79

Exhibit 1-2 Stream Table, Liquid Solvent Base Case

	1	2	3	4	5	6	7	8	9	10
		FLUE GAS	CLEAN	LEAN	RICH	RICH	LEAN SOLV	CO2 TO	STEAM	HI PRESS
From	FEED	TO	FLUE GAS	SOLVENT	SOLVENT	SOLVENT	FROM	COMPR	TO	CO2
To		ABSORBER		TO COOLER	FRM ABS	TO DESORB	DESORBER		REBOILER	
Component Mole Fraction										
H2O	0.1725	0.0727	0.0838	0.0000	0.0000	0.0000	0.0000	0.0000	1.0000	0.0000
N2	0.6644	0.7447	0.8586	0.0000	0.0000	0.0000	0.0000	0.0000	0	0.0000
O2	0.0234	0.0262	0.0302	0.0000	0.0000	0.0000	0.0000	0.0000	0	0.0000
H2S	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0	0.0000
AR	0.0080	0.0090	0.0103	0.0000	0.0000	0.0000	0.0000	0.0000	0	0.0000
SO2	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0.0000	0	0.0000
NO	0	0	0	0.0000	0.0000	0.0000	0.0000	0	0	0
NO2	0	0	0	0.0000	0.0000	0.0000	0.0000	0	0	0
GAP1X	0	0	0	0.2513	0.0252	0.0252	0.2513	0	0	0
GAP1CRBX	0	0	0	0.1482	0.3744	0.3744	0.1482	0	0	0
CO2	0.1317	0.147424	0.016998	0.0000	0.0000	0.0000	0.0000	1	0	1
TEG	0	0	0	0.6004	0.6004	0.6004	0.6004	0	0	0
Total	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000	1.0000
Mole Flow, LBMOLE/HR	1.79E+05	1.60E+05	1.39E+05	9.37E+04	9.37E+04	9.37E+04	9.37E+04	2.12E+04	6.76E+04	2.12E+04
Mass Flow, LB/HR	5.12E+06	4.77E+06	3.84E+06	2.11E+07	2.21E+07	2.21E+07	2.11E+07	9.33E+05	1.22E+06	9.33E+05
Volume Flow, CUFT/HR	7.77E+07	6.58E+07	6.45E+07	2.36E+05	2.58E+05	2.71E+05	2.50E+05	2.04E+06	5.14E+06	2.29E+04
Temperature, Deg F	135.0	104.0	178.1	198.5	178.1	266.9	290.0	105.0	743.3	124.0
Pressure, PSIA	14.70	14.70	14.70	63.00	14.70	93.00	63.00	63.00	167.70	2215.00
Vapor Fraction	1.0	1.0	1.0	0.0	0.0	0.0	0.0	1.0	1.0	1.0
Mass Enthalpy, BTU/LB	-1391.8	-1082.0	-392.8	-1634.6	-1774.8	-1725.5	-1583.4	-3838.4	-5466.6	-3923.8
Mass Density, LB/CUFT	0.066	0.073	0.059	89.412	85.643	81.521	84.611	0.458	0.237	40.748
Average Molecular Weight	28.58	29.86	27.69	225.63	235.59	235.59	225.63	44.01	18.02	44.01

The MEA equipment list is summarized in the following table:

Equipment Number	Description	Type	Design Conditions	Operating Quantity	Spares
E001	Steam Reboiler	Shell and tube SS shell; SS tubes	Duty = 1614 MMBtu/hr (1)	4	0
E002	CO2 Cooler	Shell and tube CS shell; CS tubes	Duty = 424 MMBtu/hr (1)	4	0
E003	Lean/Rich Heat Exchanger	Shell and tube CS shell; SS tubes	Duty = 1659 MMBtu/hr (1)	4	0
E004	Lean Solvent Cooler	Shell and tube CS shell; CS tubes	Duty = 624 MMBtu/hr (1)	4	0
TK001	Direct Contact Feed Cooler	Heat Exchanger Tower, CS	66 ft ID x 10 ft ht	1	0
TK002	Solvent Feed Holding Tank	Vertical cylindrical, CS	352,500 gallon capacity	1	0
P001	Rich Solvent Pump	Centrifugal, CS	9180 gpm @ 78 psi head	4	4
P002	Lean Solvent Pump	Centrifugal, CS	8810 gpm @ 74 psi head	4	4
T001	Absorber	Packed tower: CS shell; SS internals	40 ft ID x 95 ft T/T	4	0
T002	Desorber	Packed Tower: CS shell; SS internals	20 ft ID x 100 ft T/T	4	0
CPP001	CO2 Compressor Package	Integrally geared, multi-stage centrifugal, CS	953,000 lb/hr @ 2215 psia	1	0

Note 1: Total duty for four trains.

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

Equipment Number	Description	Type	Design Conditions	Operating Quantity	Spares
E001	Steam Reboiler	Shell and tube SS shell; SS tubes	Duty = 1043 MMBtu/hr (1)	4	0
E002	CO2 Cooler	Shell and tube CS shell; CS tubes	Duty = 38 MMBtu/hr (1)	4	0
E003	Lean/Rich Heat Exchanger	Shell and tube CS shell; SS tubes	Duty = 1082 MMBtu/hr (1)	4	0
E004	Lean Solvent Cooler	Shell and tube CS shell; CS tubes	Duty = 1068 MMBtu/hr (1)	4	0
TK001	Direct Contact Feed Cooler	Heat Exchanger Tower, CS	66 ft ID x 10 ft ht	1	0
TK002	Solvent Feed Holding Tank	Vertical cylindrical, CS	280,500 gallon capacity	1	0
P001	Rich Solvent Pump	Centrifugal, CS	8034 gpm @ 78 psi head	4	4
P002	Lean Solvent Pump	Centrifugal, CS	7013 gpm @ 74 psi head	4	4
T001	Absorber	Packed tower: CS shell; SS internals	35 ft ID x 95 ft T/T	4	0
T002	Desorber	Tray Tower (20 trays) CS shell and SS internals	14 ft ID x 77 ft T/T	4	0
CPP001	CO2 Compressor Package	Integrally geared, multi-stage centrifugal	953,000 lb/hr @ 2215 psia	1	0
Note 1:	Total duty for four trains.				

Cost Estimates

As mentioned above, cost estimates were conducted using the Aspen^{TM} Cost Estimator using the results from the Aspen Plus^{TM} models.

The overall heat transfer coefficients were estimated as explained in the Section: Sensitivity Analysis. Aspen Plus^{TM} was used to estimate the required heat exchange area. The heat transfer coefficients were initially estimated using approximate values published in “Process Heat Transfer” by D.Q. Kern. This resulted in a low value for the GAP-1/TEG case leading to high heat exchange area for the lean/rich solvent heat exchanger which had a disproportionate impact on overall project costs. For that reason, GE GRC did a detailed analysis of the expected overall heat transfer coefficient (HTC) for the lean/rich heat exchanger and estimated values for both the GAP-1/TEG and MEA systems. The values for this study were calculated to be 75 (GAP-1/TEG) and 93 (MEA) Btu/hr/ft^2/°F, respectively. It was also found that GPSA (Gas Processors Suppliers Association) listed an overall heat transfer coefficient of 120 to 130 Btu/hr/ft^2/°F for MEA. This would suggest that the GE generated heat transfer coefficients for both solvents may still be conservatively low but are higher than the estimates from Kern. For the lean/rich exchanger in GAP-1/TEG service, the GE calculated value of 75 Btu/hr/ft^2/°F was used. For MEA, the higher GPSA heat transfer coefficient of 120 Btu/hr/ft^2/°F was used. Actual experiments would need to be performed to confirm these results, especially for GAP-1/TEG.

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

For GAP-1/TEG, the absorber was sized as a packed tower and the desorber was sized as a trayed tower. 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 exchanger and the steam reboiler in the desorber column.

For GAP-1/TEG two cases were run. In the first case, the metallurgy for the carbon separation unit was assumed to be carbon steel with the exception of heat exchanger tubes in hot rich solvent service and column internals (trays or packing), which were made of stainless steel. In the second case, all parts were assumed to be carbon steel.

GAP-1/TEG System (Carbon Steel/Stainless Steel Option)

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 cooperative agreement, 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 cooperative agreement is:

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

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

$$= 0.79 \times \$484.5 \text{ Million} \sim \$383 \text{ Million}$$

The relative capital cost of amino-silicone solvent is 85% of the conventional MEA case (based on modeling performed by GE in this cooperative agreement, using commercially available Aspen™ cost estimation software). Compared to the DOE/NETL-2007/1281 report capacity adjusted CAPEX of $383 Million for the conventional MEA system, the CAPEX for amino-silicone would be $326 Million[1].

GAP-1/TEG System (Carbon Steel Option)

To check the sensitivity of the cost to the metallurgy requirement, the cost of the GAP-1/TEG system was repeated assuming a lower metallurgy for the tower internals and heat exchanger tubes (substituting carbon steel for stainless steel). Testing has shown that the GAP-1/TEG 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. These tests will be conducted in Phase II of the award.

Using the same cost calculation methodology as above, if only carbon steel is required, the overall unit costs for amino silicone would fall to ~57% of the MEA case, or $218 Million[1].

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	3,372
CO_2 Compression	41,885
Cooling Water Fans/Pumps	9,965
TOTAL AUXILIARIES, kWe	64,476
COOLING WATER, ton/hr	40,254
STEAM, ton/hr	759.7

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[1] These estimates are based on commercial ASPEN cost estimator software and scaling based on DOE/NETL-2007/1281 report.DOE/NETL-2007/1281 report.

The system utilities for a typical amino-silicone system are summarized in the following table:

POWER SUMMARY	
AUXILIARY LOAD SUMMARY, kWe	
Feed Gas Blower	9,254
CO_2 Separation Auxiliaries	2,696
CO_2 Compression	30,261
Cooling Water Fans/Pumps	11,264
TOTAL AUXILIARIES, kWe	53,475
COOLING WATER, ton/hr	45,500
STEAM, ton/hr	491.0

The water falls for power plant net efficiency and energy penalty due to CCS are shown in Figure 4 and Figure 5. The 1st column is for the MEA solvent. The remaining columns are for amino-silicone solvent system. The 2nd column is for a desorber temperature of 143 °C, cross heat exchanger temperature approach of 16.7 °C (30 °F) and steam extraction temperature of 395 °C (743 °F). The 3rd column is with a cross heat exchanger temperature approach of 5.6 °C (10 °F). The 4th column is with a desorber temperature of 160 °C. The 5th column is with a lower steam extraction temperature of 171 °C. The amino-silicone system has a significantly lower energy penalty when compared with MEA.

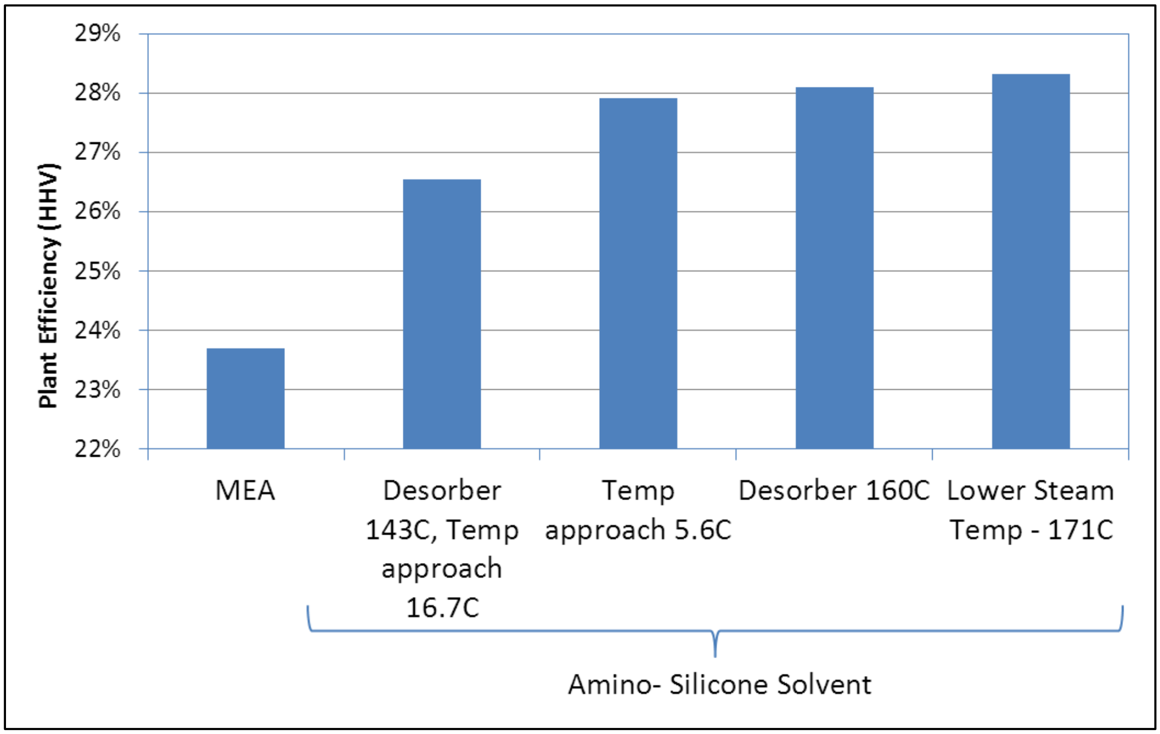

Figure 4: Plant Efficiency Based on HHV.

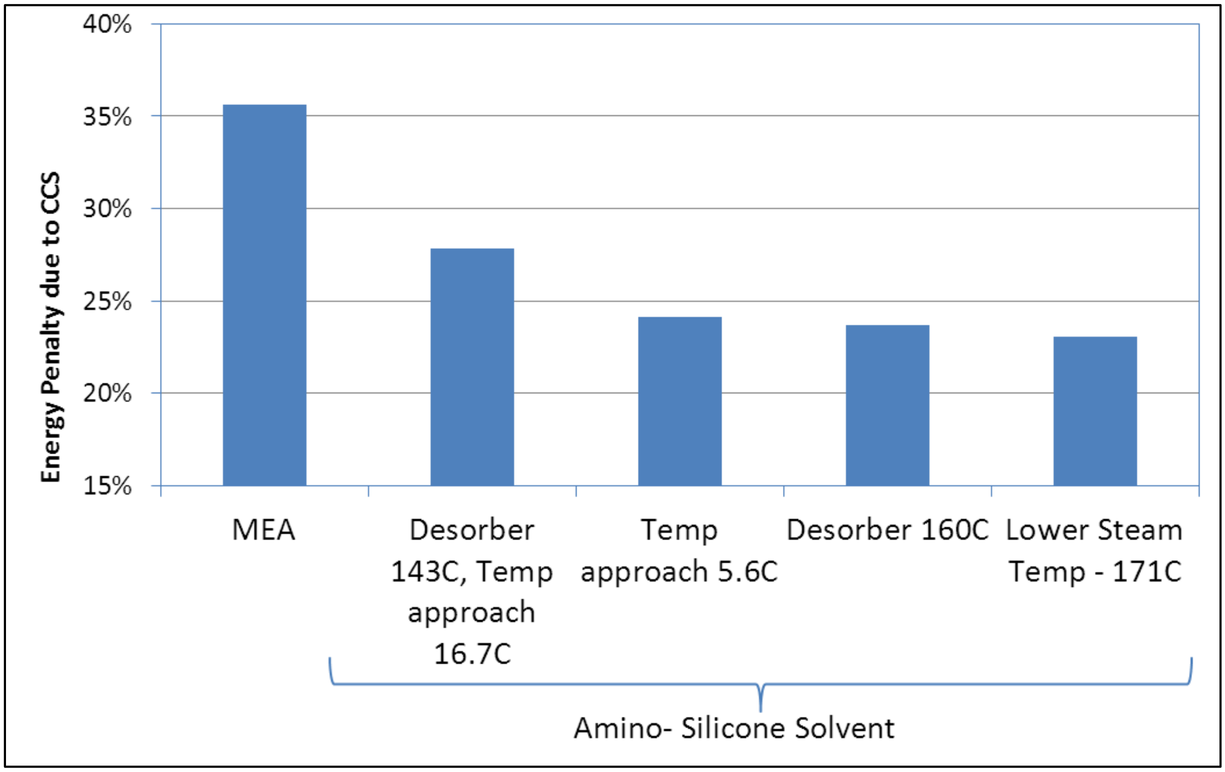

Figure 5: Energy Penalty due to CCS

Cost of Electricity

Economic Analysis Assumptions:

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 121 lbs/hr. The lab data indicated that GAP-1 thermally degrades by 14% over 90 days at 150 °C based on CO_2 pickup. This resulted in a GAP-1 loss of 24 lbs/hr. So, the total GAP-1 loss is 145 lbs/hr, due to cumulative effects of SO_2 degradation and thermal degradation.

Based on data from DOE/NETL-2007/1281 report adjusted to DOE specified assumptions for this study using commercially available Aspen™ cost estimation software, the annual O&M costs for MEA are estimated to be $354 Million. Using the solvent cost estimated by GE contractor SiVance, the O&M costs for GAP-1/TEG are estimated to be 5% higher than that of MEA. Therefore, the annual O&M costs for GAP-1/TEG solvent are estimated to be $372 Million[2].

The water fall chart for increase in cost of electricity (COE) over a non-capture case is shown in Figure 6, using the following equation which is specified in the cooperative agreement and is marked "simple" in the chart":

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[2] These estimates are based on commercial ASPEN software and scaling based on DOE/NETL-2007/1281 report.

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

The 1st column is for the MEA solvent. The remaining columns are for the amino-silicone solvent system. The 2nd column is for a desorber temperature of 143 °C, cross heat exchanger temperature approach of 16.7 °C, steam extraction temperature of 395 °C (743 °F) and stainless steel construction for critical components. The 3rd column is with a cross heat exchanger temperature approach of 5.6 °C. The 4th column is with a desorber temperature of 160 °C. The 5th column is with carbon steel construction. The 6th column is with a lower steam extraction temperature of 171 °C. The amino-silicone system has significantly lower COE when compared with MEA.

Using the DOE assumptions listed above, the increase in COE, over a plant without CO_2 capture, is shown in Figure 7. This figure is marked "detailed" since more detailed calculations were performed, while Figure 6 is marked "simple" since the above simple equation was used.

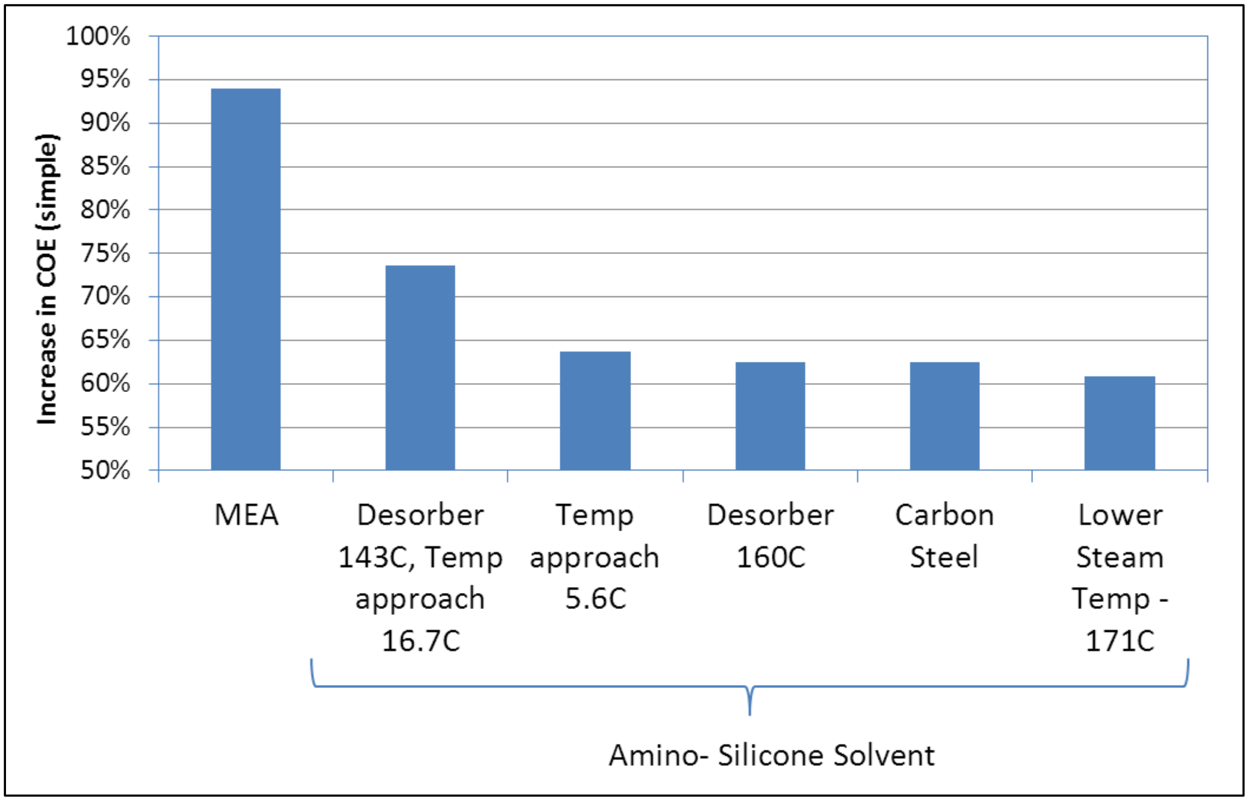

Figure 6: Increase in COE using simplified calculation as set in the cooperative agreement.

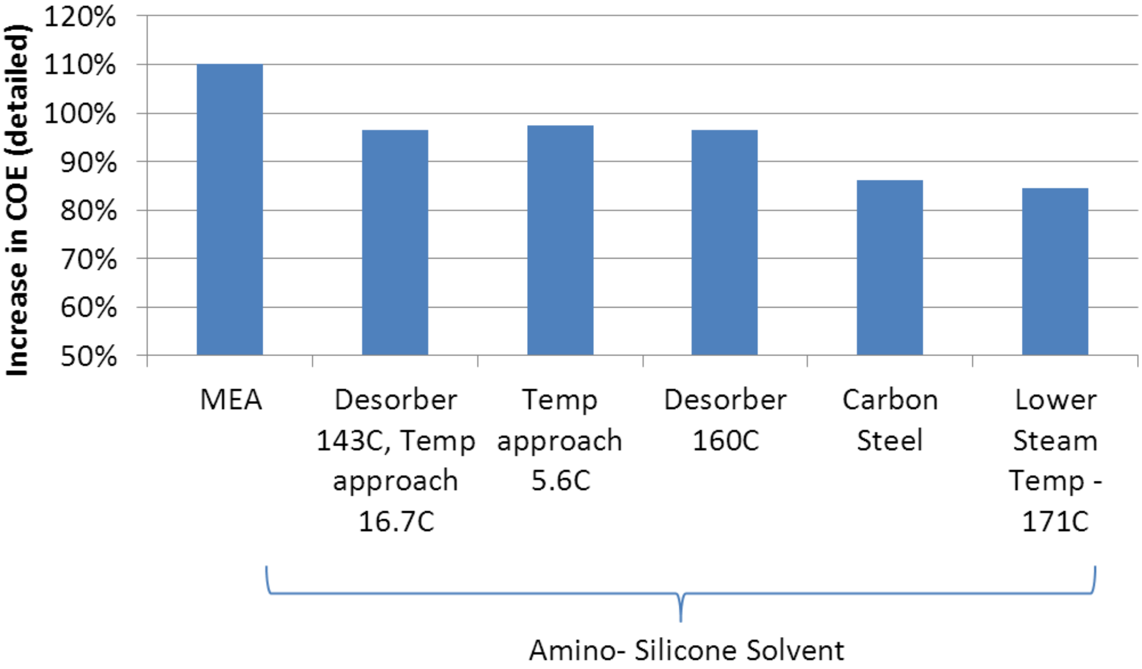

Figure 7: Increase in COE as calculated from detailed analysis using energy penalty and Capex estimates.

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.

	MEA	Amino-Silicone Solvent
Cost of Steam - $/1000 lbs	**$ 11.31**	**$ 9.47**
Cost of CO2 - $/ton	**$ 26.64**	**$ 31.73**
Removal cost for CO2 - $/ton	**$ 53.51**	**$ 49.59**
Avoided Cost for CO2 - $/ton	**$ 88.42**	**$ 67.17**
% decrease in PC Plant Efficiency	**13.1%**	**8.7%**

Sensitivity Analysis

Sensitivity analysis was carried out for the model for 1) GAP-1/TEG heat transfer coefficient, 2) temperature of desorber, and 3) temperature approach in the rich/lean heat exchanger. The models used for the analysis are explained in the previous sections.

Heat Transfer Coefficient

The overall heat transfer coefficient for shell and tube heat exchangers can be calculated from the following expression[i]:

$$U_o = \frac{1}{1/h_o + R_{do} + xA_o/k_w A_{wm} + (1/h_i + R_{di})A_o/A_i} \qquad \text{Equation 1}$$

where h_0 and h_i are individual film heat-transfer coefficients, R_{do} and R_{di} are fouling resistances; and $(xA_0/k_w A_{wm})$ is wall resistance.

Two separate methods were used to calculate individual film heat-transfer coefficients for tube and shell sides.

Shell-Side Individual Film Heat-Transfer Coefficient

Shell-side heat-transfer coefficient for an ideal tube bank h_k can be determined from the following expression[i]:

$$h_k = j_k c \frac{W}{S_m}\left(\frac{k}{c\mu}\right)^{2/3}\left(\frac{\mu_b}{\mu_w}\right)^{0.14} \qquad \text{Equation 2}$$

where j_k is the factor determined from the correlation for j-factor for ideal tube bank (Figure 8), c is specific heat, k is the thermal conductivity, μ_b is bulk viscosity of the solvent, μ_w is viscosity evaluated at the mean surface temperature, W is mass flow rate, and S_m is one cross-flow section S_m.

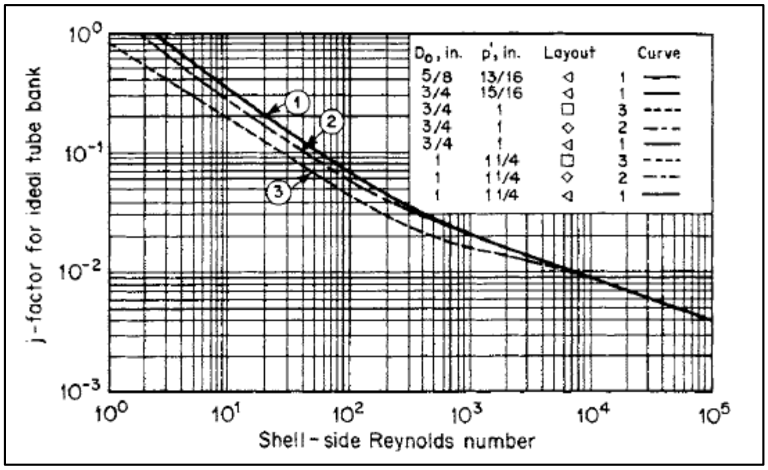

Figure 8: Correlation of j factor for ideal tube bank[i].

The shell side Reynolds number can be determined from the following equation[i]:

$$(N_{\mathrm{Re}})_s = D_o W/\mu_b S_m \qquad \text{Equation 3}$$

Steps for calculation of shell-side heat transfer coefficient are described below:

1) Identify assumptions for these calculations:
    a. Reynolds number on the shell side for MEA and GAP-1/TEG system is the same, and it equals 1,000.
    b. Ratio of bulk to wall viscosity is assumed to be 10. Due to the higher temperature of the wall surface versus bulk, the viscosity will be lower at the surface. Estimated value has little impact on heat transfer coefficient due to small exponent in the Equation 2. Ten was chosen as a conservative estimate.
    c. Tube diameter is 1.5 inch.

2) Calculate ratio of W/S_m from Equation 3.
3) Find j_k value from the plot in Figure 8.
4) Substitute physical properties of the solvent, value of j_k, and the value of W/S_m into the equation 2 to find h_o.

Tube-Side Individual Film Heat-Transfer Coefficient

Tube side heat transfer coefficient for circular tube can be determined from the following Nusselt number correlation for laminar flow[ii]:

$$Nu = 1.86 * (Re * \mathrm{Pr})^{0.33} * (\frac{d}{L})^{0.33} (\frac{\mu_{wall}}{\mu_{bulk}})^{0.14} \qquad \text{Equation 4}$$

Also, Nusselt number can be correlated to the heat transfer coefficient h through the following expression[iii]:

$$Nu = \frac{h * d}{12k} \qquad \text{Equation 5}$$

Below are the steps for the calculation of tube-side heat transfer coefficient:

1) Identify assumption for these calculations:
    a. Reynolds number on the tube side for MEA and GAP-1/TEG system is the same, and it equals 1,000 (laminar flow). This number was selected as a moderate value corresponding to a flow of ~ 10 ft/sec.
    b. Tube diameter is 1.5 in.
2) Calculate Prandtl number for each solvent system.
3) Calculate Nusselt number from the Equation 4.
4) Calculate h_i from the Equation 5.

Overall Heat-Transfer Coefficient

Overall heat transfer coefficient can be calculated from Equation 1, and below are the steps for calculations:

1) Identify assumptions for these calculations:
    a. Thickness of the pipe wall is 0.25 in.
    b. Pipe material is carbon steel.
    c. Fouling coefficient is 5,000 W/m^2*K.

2) Use Equation 1 to determine overall heat transfer coefficient U.

These calculations were used to determine the overall heat transfer coefficients for 30/70 MEA/water system and compare it to the 60/40 GAP-1/TEG system. The values of overall heat transfer coefficient for 60/40 GAP-1/TEG and 30/70 MEA/water are 75 and 93 BTU/(hr*ft^2*F), respectively.

It has to be noted that this value of U is specific to the assumptions made and considered conditions. Due to the high viscosity of the rich GAP-1/TEG solvent, turbulent flow might be a challenge for the tube side of the heat exchanger, and pressure drop would also need to be considered for the final design. So, a velocity in the laminar regime was chosen. To increase the overall heat transfer, shell side Reynolds number can be potentially increased.

Sensitivity to Heat Transfer Coefficient

The heat transfer coefficient for GAP-1 was varied from 25-100 Btu/(hr ft^2 °F). The range was selected based on estimates for heat transfer coefficients calculated as explained in the previous section. The heat transfer area for the rich/lean heat exchanger and the lean cooler was calculated using Aspen Plus^{TM}. The results for the calculated area are shown in Figure 9.

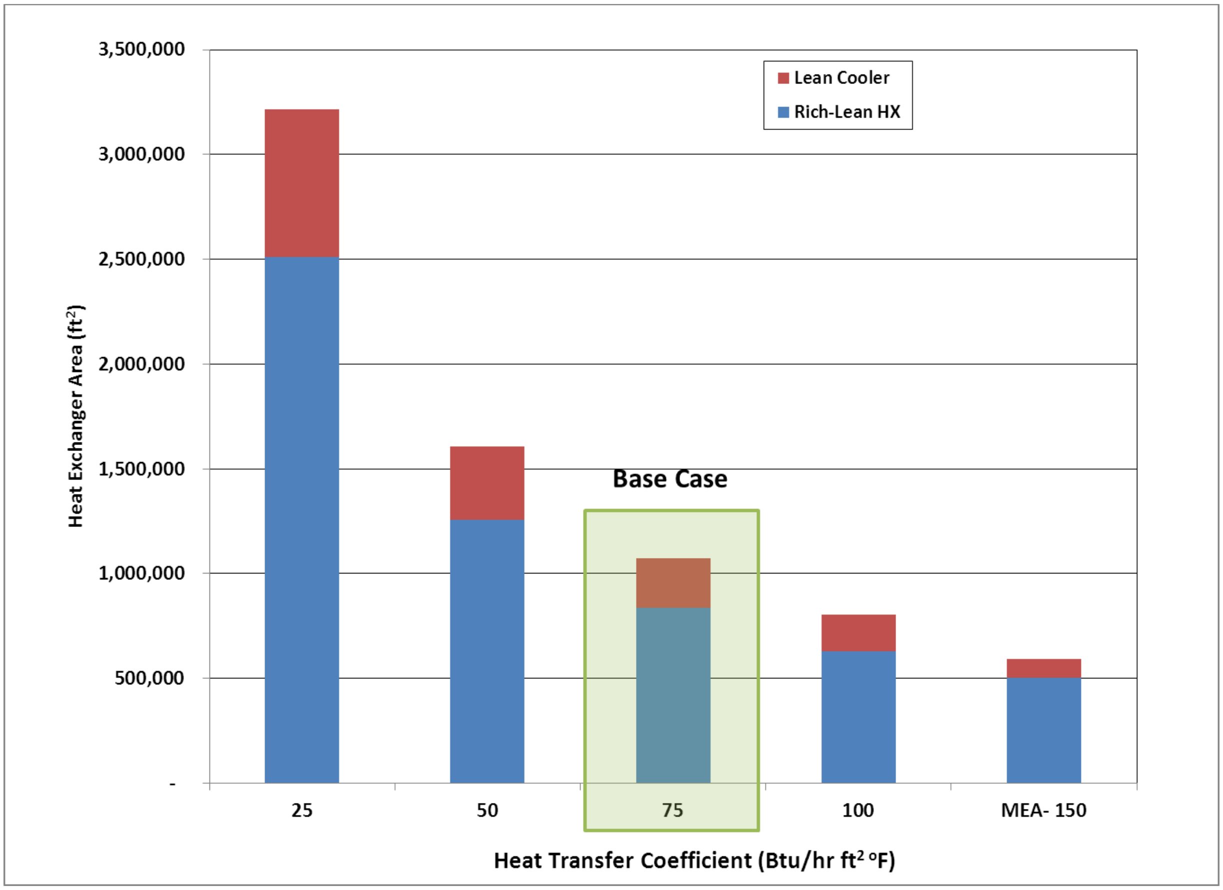

Figure 9: Heat exchanger area dependence on heat transfer coefficient of GAP-1/TEG vs. MEA.

Aspen Cost Estimator model was used to estimate the cost for the two heat exchangers as explained in cost estimation section. The sensitivity analysis was carried out to estimate the cost variation with heat transfer coefficient. The costs of the GAP-1/TEG lean cooler relative to MEA are shown in Figure 10. The costs of the GAP-1/TEG rich-lean heat exchanger relative to MEA are shown in Figure 11. The smaller bars are the equipment cost and the large bars include the cost of equipment, piping, instrumentation, installation and paint etc.

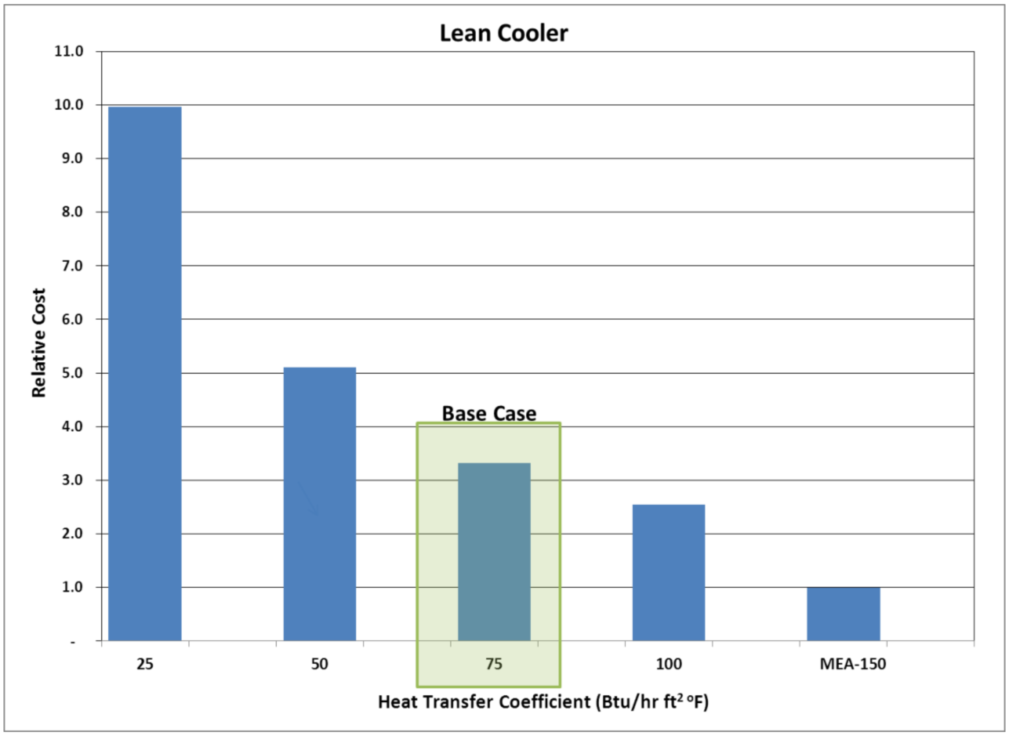

Figure 10: Heat exchanger cost dependence on heat transfer coefficient of GAP-1/TEG vs. MEA.

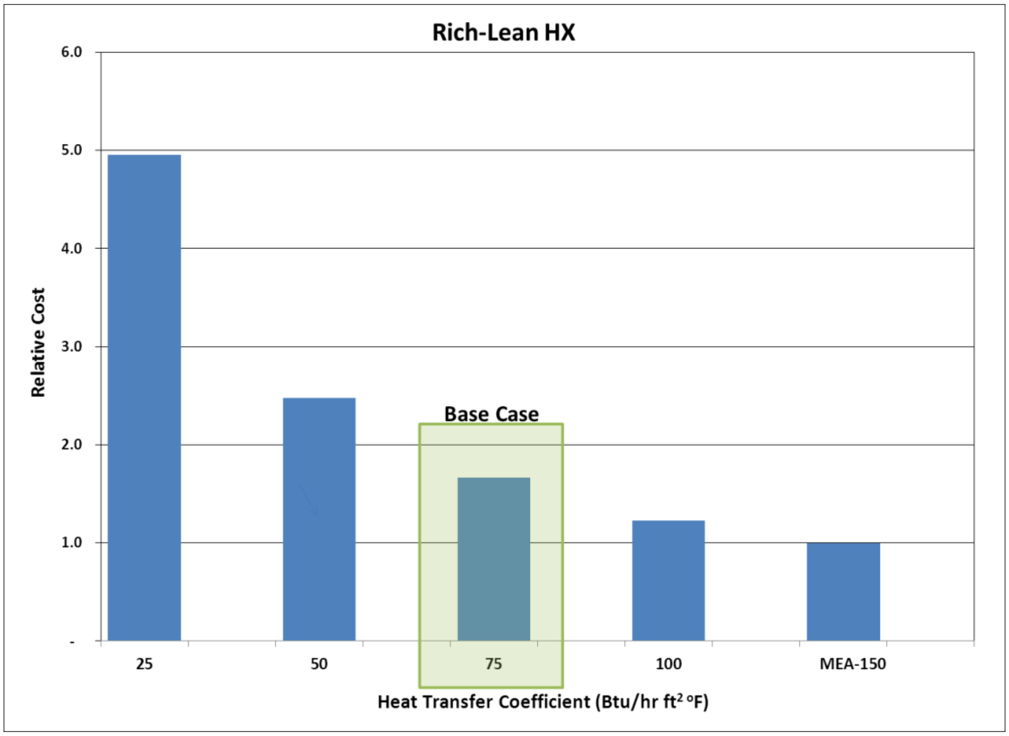

Figure 11: Heat exchanger cost dependence on heat transfer coefficient of GAP-1/TEG vs. MEA.

It is clear from Figures 9- 11 that the heat transfer coefficient of the solvent has a large impact on the heat exchanger area and the cost for the two components. The variation of total Capex for the overall system relative to MEA due to the heat transfer coefficient is shown in Figure 12. As shown, the heat transfer coefficient of the GAP-1/TEG solvent is a big factor in determining the overall cost of the process.

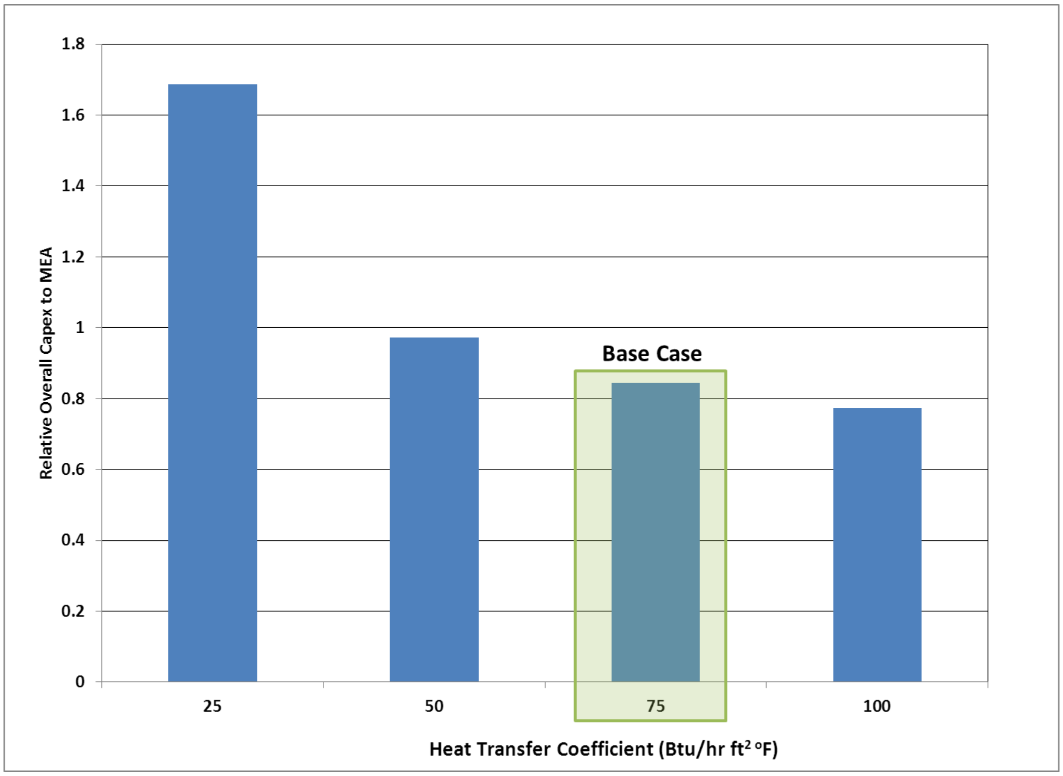

Figure 12: Overall Capex dependence on heat transfer coefficient of GAP-1/TEG.

Sensitivity to Desorption Temperature

Another set of sensitivity analysis was carried out for overall system Capex with respect to desorption temperature. The HT coefficient used for this system analysis was fixed at 75 Btu/(hr ft^2 °F). It should be noted that when the desorption temperature is changed then other variables such as the desorption pressure and CO_2 in the lean solvent are affected. Hence, the total solvent flow rate changes. These changes are summarized in the following table.

Temperature (°C)	Pressure (psi)	CO_2 in solvent after desorber
120	36	4.5%
143	63	2.9%
160	75	2.0%
180	85	1.8%

The effect of desorber temperature on energy penalty due to CCS for a steam temperature of 743 °F is shown in Figure 13.

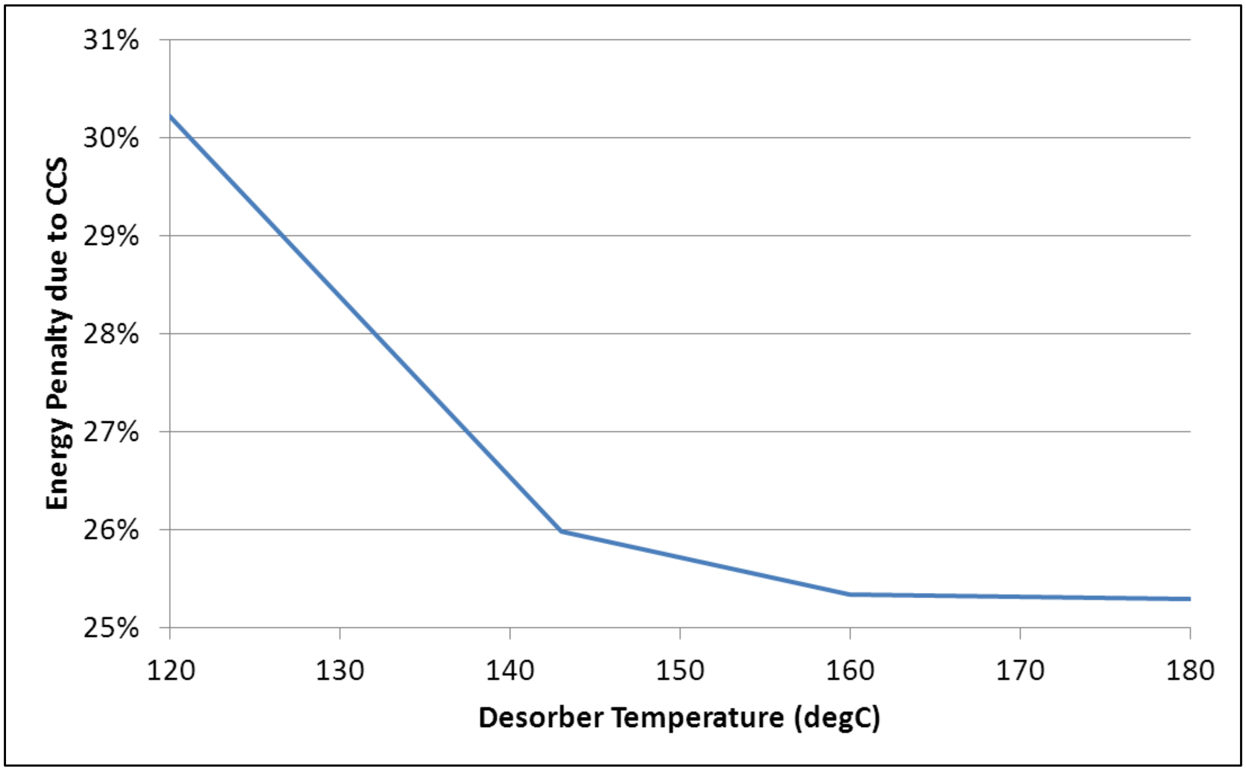

Figure 13: Effect of desorber temperature on energy penalty of CCS using GAP-1/TEG as a solvent.

The Aspen Plus model was used to size the equipment and then Aspen Plus Cost Estimator was used to estimate the overall Capex for the system as explained in the previous sections. The effect of desorption temperature on Capex for the CO_2 separation unit based on GAP-1/TEG relative to MEA is shown in Figure 14.

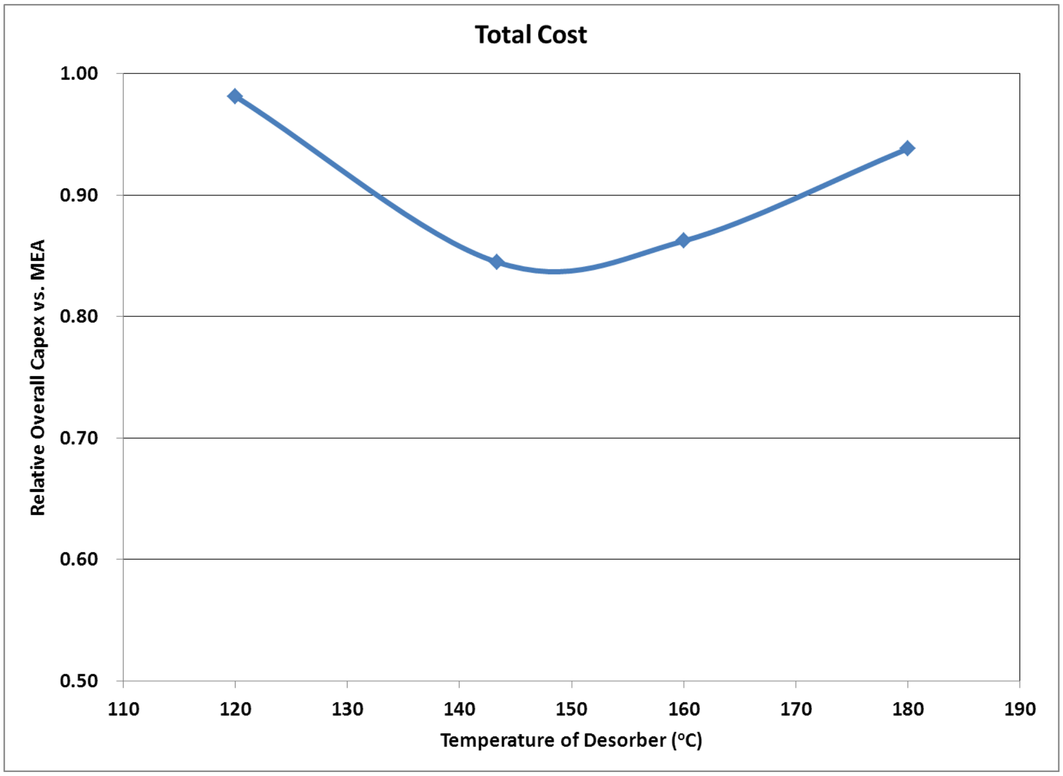

Figure 14: Overall Capex for the CO_2 separation system using GAP-1/TEG solvent vs. desorption temperature.

It is clear from Figure 14 that the overall Capex for the system first goes down with increasing temperature because a lower solvent flow rate is required due to lower CO_2 concentration in the lean solvent after the desorber based on the desorption isotherms. After a certain point the overall Capex of the system increases with desorption temperature because the CO_2 in the lean solvent reaches a minimum whereas the heat transfer areas for rich/lean heat exchanger and lean cooler keep on increasing based on an increasing temperature difference between rich solvent and lean solvent (related to absorber temperature and desorber temperature).

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 35.6% and the energy penalty for amino-silicone solvent is 23.7% using a steam temperature of 395 °C (743 °F). If the steam temperature is lowered to 171 °C (340 °F), the energy penalty for the amino-silicone solvent is reduced to 23%. The increase in

COE over the non-capture case for MEA is ~110% and increase in COE for amino-silicone solvent is 86% at a steam temperature of 395 °C (743 °F). If the steam temperature is lowered to 171 °C (340 °F), the increase in COE for the amino-silicone solvent is reduced to 85%.

References:

1. Perry’s chemical engineering handbook, 8th edition.
2. “Chemical Engineering Design: Principles, practice and economics of plant and process design”, G. Towler, R. Sinnott.
3. Simplified Approach to estimating tube side heat transfer coefficients.