

Systems Analysis of Ionic Liquids Sorbents for CO₂ Capture

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ABSTRACT

Trimeric Corporation performed systems analysis studies to evaluate the potential of ionic liquids (IL) as a solvent for CO₂ capture. The systems analysis included a sensitivity study identifying key ionic liquid properties that govern parasitic power requirements and a preliminary economic evaluation to estimate the cost of electricity from a post-combustion capture coal-fired power plant (500 MW gross). The capture process was evaluated using hypothetical ionic liquids having thermodynamic and physical properties representative of properties for ionic liquids modeled and synthesized at the University of Notre Dame. Conventional absorption-stripping and thermal swing adsorption (TSA) processes were both examined; results showed the absorption-stripping process to be the more economical approach for CO₂ capture with IL.

The primary variable of interest for the sensitivity study was the enthalpy of reaction, with water solubility in the ionic liquid a secondary factor. The performance metric for the sensitivity study was the total equivalent work that must be supplied by the power plant to run the CO₂ capture process. Within the parameters of the study, a minimum equivalent work was determined for an ionic liquid with a heat of reaction of 50 kJ/mol and partial miscibility with water. A lower heat of reaction led to higher energy penalties for sensible heating and solvent pumping, while a higher heat of reaction led to increased regeneration energy. Economic analysis of four different solvent regeneration conditions showed the potential for an IL-based capture process to have a cost of electricity 10-17% less than with a conventional MEA-based process.

INTRODUCTION/BACKGROUND

The complete IL systems analysis included: 1) a process modeling sensitivity study to identify the key IL properties that govern the parasitic power requirements and 2) a preliminary economic evaluation to estimate the cost of electricity from a coal-fired power plant with post-combustion CO₂ capture using ionic liquids at different operating conditions. The process economics were then compared to a base case monoethanolamine (MEA) unit, since amine-based scrubbing is one of the most likely near-term options for post-combustion CO₂ capture. MEA unit costs were obtained from a previous DOE/NETL study completed by Trimeric in 2007¹. The sensitivity study and preliminary economic analysis were conducted for a hypothetical ionic liquid having thermodynamic and physical properties representative of the properties for ionic liquids modeled and synthesized by Notre Dame.

RESULTS SUMMARY

Trimeric conducted a sensitivity study using the HYSYS process simulation software for a hypothetical ionic liquid CO₂-capture process. The goal for the sensitivity analysis was to provide guidance to the developers of IL compounds about target properties for CO₂ capture. The primary variable of interest for the sensitivity study was the enthalpy of reaction (ΔH°_{rxn}) between IL and CO₂, with water solubility in IL as a secondary variable. The performance metric for the sensitivity study was the total equivalent work that must be supplied by the power plant to run the CO₂ capture process, which would result in a derating of the power plant electricity output.

Within the range covered by the sensitivity analysis (ΔH°_{rxn} of 40 to 60 kJ/mol, partially and fully miscible with water, reference state of vapor phase CO₂ at standard conditions of 25 °C and 1 bar), the minimum equivalent work (87.2 MW) was with an ionic liquid having a 50 kJ/mol heat of reaction with CO₂ and partial miscibility with water. A minimum exists because of offsetting effects of CO₂ loading and regeneration energy. At ΔH°_{rxn} less than 50 kJ/mol, the capacity of the IL is reduced, leading to relatively high energy penalties for sensible heating and pumping of the liquid solvent. At ΔH°_{rxn} greater than 50 kJ/mol, the increased regeneration energy becomes the primary penalty.

The total equivalent work was extremely sensitive to changes in ΔH°_{rxn} between 40 and 50 kJ/mol, with the 40 kJ/mol minimum equivalent work about 70% higher than the 50 kJ/mol minimum equivalent work. The minimum equivalent work for the 60 kJ/mol case was only 3% higher than the 50 kJ/mol minimum equivalent work, but the 60 kJ/mol minimum occurred at a much higher regeneration temperature and pressure, which affects the thermal stability of the IL. The equivalent work analysis was based on a 1:1 IL:CO₂ reaction stoichiometry.

The total equivalent work for the 50 kJ/mol cases was not particularly sensitive to regeneration temperature at a given regeneration pressure, and the equivalent work for a wide range of regeneration temperatures and pressures varied only by about 10-15%. The water miscibility had a secondary effect, with fully-miscible systems resulting in slightly higher (3-5%) total equivalent work than partially miscible systems.

Detailed modeling and economic analysis were conducted for four IL cases operating with different stripper temperatures and pressures. These cases are listed below.

- Case 1: Low stripper temperature and pressure (149°C and 136 kPa; 300°F and 5 psig)
- Case 2: Mid stripper temperature and pressure (166°C and 584 kPa; 330°F and 70 psig)
- Case 3: High stripper temperature and mid pressure (199°C and 584 kPa; 390°F and 70 psig)
- Case 4: High stripper temperature and pressure (204°C and 791 kPa; 400°F and 100 psig)

Cases 1 and 2 were lower-temperature regeneration cases, since there is uncertainty about long-term thermal stability of some functionalized IL compounds. Cases 3 and 4 were increased regeneration temperature and pressure cases to evaluate the potential for reduction in downstream CO₂ compression costs.

Rigorous process simulations with mass and energy balances were prepared. Then, the equipment in the process was sized and selected, and purchased equipment costs were developed from vendor information or costing software. Finally, capital costs, operating costs, incremental cost of electricity, and cost of CO₂ emissions were estimated. Costs for the four IL cases were compared to a baseline MEA case taken from a previous DOE SBIR¹.

The design basis for these evaluations was a 500 MW gross conventional coal-fired power plant using Illinois #6 subbituminous coal. A wet flue gas desulfurization (FGD) unit was assumed to be located upstream of the CO₂ capture unit. The target CO₂ removal was 90%. Any captured CO₂ was assumed to be compressed to pipeline pressure (15.2 MPa; 2200 psia). The entire CO₂ capture system consisted of a single inlet gas train, multiple parallel ionic liquid units, and a single, common CO₂ compression train.

Trimeric estimated that CO₂ capture increased the base cost of electricity from 5 ¢/kWh to 10.6 ¢/kWh for MEA and 8.8 to 9.5 ¢/kWh for the hypothetical ionic liquids, excluding the solvent management costs associated with degradation or other losses. The corresponding cost per metric tonne CO₂ avoided was \$66.8/tonne CO₂ and \$44.8 to 52.5/tonne CO₂, respectively. Derated capacities were 339 MW_e for MEA and 361 to 368 MW_e for the ionic liquids. When compared to the base case, the ionic liquids reduced the reboiler steam requirements by 41 to 48%, reduced derating due to CO₂ capture by 13 to 18%, and reduced the cost of CO₂ avoided by 21 to 33%.

The results are encouraging in that they show that the ionic liquids can potentially offer significant cost savings over the base MEA solvent. However, the economic analysis should be considered preliminary given the early phase of the IL development. Different ionic liquids are still being identified and tested,

and, at the time of the systems analysis, no ionic liquid existed that had all of the optimal properties used in the model. In addition, there are several key parameters that need to be explored more thoroughly in the future that could significantly impact the economics of the IL solutions. The most important is the degradation rate of the IL. Current data shows a wide range of possible degradation rates depending on the specific ionic liquid. Because of this uncertainty, degradation was not included for either MEA or the IL in the economic comparisons reported in this study.

Other important variables that could impact the ionic liquid economics are the material of construction (assumed similar to MEA until corrosion tests are completed), the kinetic rate for CO₂ absorption that affects packing heights in the absorber and stripper (assumed similar to MEA until tests are completed), and other equipment features (i.e., low design temperature differential in the rich/lean exchanger and IL solution reclamation).

DESIGN BASIS/PROCESS DESCRIPTION

Two process configurations were chosen for a complete systems analysis: a conventional absorption-stripping process and a thermal swing adsorption (TSA) process configuration. The absorption-stripping process is similar to conventional amine CO₂ capture processes, while the TSA process uses IL supported on silica gel packing to adsorb CO₂, with regeneration using high pressure steam.

CONVENTIONAL ABSORPTION-STRIPPING

A generic process flow diagram for an IL-based CO₂ capture system is shown in Figure 1. The inlet flue gas passes through a blower to increase the pressure to 111.0 kPa (16.1 psia) and a direct contact cooler to lower the temperature to 40°C (104°F). The gas then enters the bottom of the absorber. Cooled, lean IL solution enters the top of the absorber, and rich solution exits the bottom of the absorber. Flue gas exits from the top of the absorber and flows to the stack. Rich solution exchanges heat with hot lean solution. The preheated rich solution flows to the stripper where CO₂ desorbs from solution. Warm stripper overheads flows to a condenser where the vapor is cooled and water is condensed. The remaining CO₂ vapor then flows to compression. A steam-heated reboiler provides heat to the stripper column for CO₂ desorption and sensible heating of the liquid. Hot lean solution exits from the bottom of the stripper and is cooled through cross exchange with the rich solution.

The captured CO₂ is compressed to pipeline pressure of 15.27 MPa (2200 psig). The compressor train has multiple stages of centrifugal compression with interstage coolers and separators that increase the

pressure to about 8.4 MPa (1200 psig). An electric submersible pump (ESP) then takes the CO₂ to its final pipeline pressure. A dehydration unit is included between the final compression stage and the ESP to dry the CO₂. Minor differences in the equipment required by the conventional amine and ionic liquid processes are:

- 1) The lean solution pump may not be required in the ionic liquid process if the stripper is operated at elevated pressure (>586 kPa or 85 psia, provided by the upstream rich solution pump);
- 2) An absorber intercooler is used with the ionic liquid process to reduce the effects of the temperature bulge in the absorber. The high water content in the MEA solution acts as a heat sink for the exothermic amine-CO₂ reaction, and lessens the effect of the temperature bulge; the absence of water in the IL process necessitates the intercooling in the absorber; and
- 3) Fewer compression stages may be used with the ionic liquids process if the stripper operates at higher pressure (MEA strippers typically operate in the 35-69 kPa or 5-10 psig range).

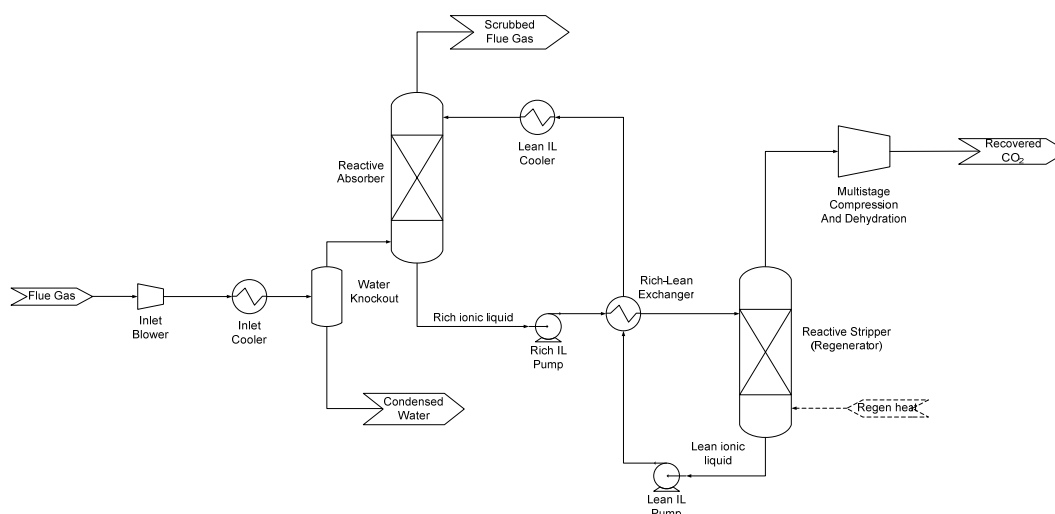


Figure 1. IL Absorption-Stripping CO₂ Capture Process Flow Diagram.

THERMAL SWING ADSORPTION (TSA)

Early IL solvents had extremely high viscosities (>1000 cP) when complexed with CO₂, so alternative CO₂ capture process configurations, including TSA, were considered as replacements for the standard absorption-stripping process. Subsequent R&D yielded IL compounds with viscosities that are compatible with absorption-stripping processes, but the TSA system analysis is included in this paper for completeness. A simplified process flow diagram of a hypothetical TSA process is shown in Figure 2. The flue gas adsorption bed is a vessel packed with tubes that contain IL sorbent supported on solid silica gel pellets. CO₂ from the cooled flue gas reacts with the IL sorbent as it passes through the

adsorption bed. The cleaned flue gas exits the top of the adsorption bed and exits with the power plant stack.

Once CO₂ breakthrough occurs and the bed on the feed step is saturated with CO₂, another bed is brought on line for flue gas feed. The saturated bed is taken offline and regenerated using saturated steam at two different pressures, which can be supplied as a slipstream from the steam turbine feed. The heat of vaporization supplied by 160 psia saturated steam passing through the shell of the adsorption vessel is used to heat the adsorbent support matrix to 130°C and desorb the CO₂ from IL sorbent; the condensate exits the adsorption vessel into a condensate knockout pot, and the condensate is returned to the steam generation facility.

A slipstream of the high-pressure steam is passed through a turbine to drop the pressure to 35 psia (the energy produced from reducing the steam pressure can be utilized and partially offset the overall steam requirements). The superheated 35 psia steam is passed through the tubes and used to sweep the desorbed CO₂ from the IL. An effluent mixture of superheated steam, 97% CO₂ (dry-basis) and 3% N₂ (dry-basis) exits the bottom of the bed during the regeneration step and is sent to a compression train. Once the bed has been fully regenerated, water supplied from a cooling tower at 85°F is passed through the shell side of the adsorption bed to cool the vessel and support matrix back to flue gas temperatures. The hot cooling water exits the bottom of the absorber and is recycled back to the cooling tower. Once the bed has been cooled, feed gas can be re-introduced to start the three-step process again.

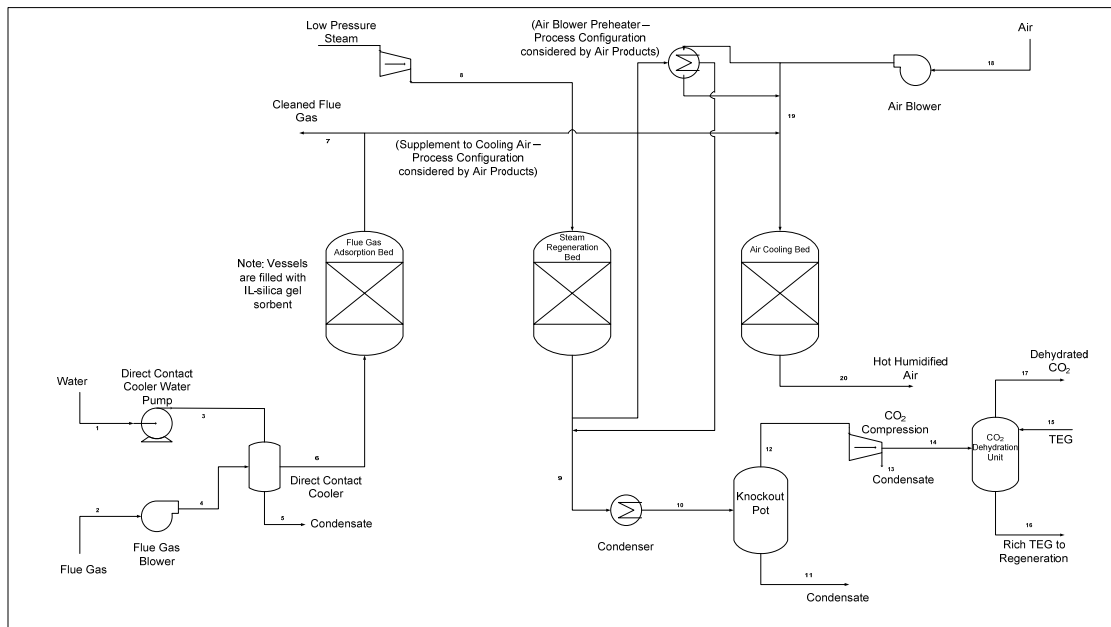


Figure 2. IL Thermal Swing Adsorption CO₂ Capture Process Flow Diagram.

IL PROPERTIES

Material and energy balances were derived from a HYSYS process model constructed using a hypothetical IL compound with properties that are representative of IL synthesized or modeled by Notre Dame but, at the time, not necessarily associated with a specific, single compound. All IL properties and reaction data were gathered from experiments and molecular modeling. The key IL properties used in the HYSYS model are listed in Table 1.

Table 1. Ionic Liquid Properties Assumed for Process Modeling.

	Value	Units
Molecular weight	500.37	g/mol
Liquid density	1.8412 - 0.001(T)	Density [=] kmol/m ³ ; T [=] K
Viscosity	100 (at 40°C)	cP
Heat capacity (no temperature dependence)	388.8	J/mol-K
Enthalpy of reaction	40, 50, and 60	kJ/mol (CO ₂ basis, vapor reference state at standard conditions)

SENSITIVITY ANALYSIS

An optimum heat of reaction exists for the model ionic liquid at approximately 50 kJ/mol. Within the range covered by the sensitivity analysis (40–60 kJ/mol), the minimum equivalent work (W_{eq}) was calculated to be 87.2 MW. A minimum equivalent work exists because of offsetting effects of CO₂ loading and regeneration energy. At a ΔH_{rxn} value less than 50 kJ/mol, the capacity of the IL is reduced, leading to relatively high energy penalties for sensible heating and pumping of the liquid solvent from the increased liquid circulation rates. At a ΔH_{rxn} greater than 50 kJ/mol, the increased regeneration energy and pumping power (from larger pressure difference between the absorber and regenerator) become the primary penalties. The equivalent work analysis was based on a 1:1 IL:CO₂ reaction stoichiometry.

The total equivalent work was extremely sensitive to changes over the range of 40 to 50 kJ/mol, with the minimum equivalent work at 40 kJ/mol about 70% higher than the minimum equivalent work at 50 kJ/mol. The minimum equivalent work for the 60 kJ/mol case was only 3% higher than the minimum equivalent work at 50 kJ/mol, but the minimum for the 60 kJ/mol case occurred at a much higher regeneration temperature and pressure – which underscores the need to have good IL thermal stability.

Research on amine-based systems indicated that under certain conditions, a higher heat of reaction always leads to a lower equivalent work.² As a result, there is no optimum heat of reaction - provided that no overstripping occurs in the regenerator. At a fixed temperature, an increase in the value of the heat of reaction leads to a higher required heat input into the regenerator; however, an increased ΔH_{rxn} value also increases the pressure in the regenerator. As a result, the power required to compress the CO₂ from regenerator pressure to pipeline pressure decreases. The benefit of the decreased compression work requirement greatly outweighs the increased regenerator thermal compression (i.e. heat input). Thus, for amine-based systems a higher heat of reaction can be favorable provided that the inherent capacity remains constant as ΔH_{rxn} changes. Any overstripping that occurs shifts the process away from equilibrium and negates the thermal compression savings.

The inherent capacity refers to the difference in equilibrium CO₂ loading of the rich solvent (at absorber bottoms conditions) compared to the CO₂ loading of the lean solvent (at absorber top conditions). At the bottom of the absorber, the CO₂ partial pressure is typically about 0.13 bar. With 90 % removal of CO₂, the CO₂ partial pressure at the top of the absorber is about 0.013 bar.

The thermodynamic properties for the ionic liquids suggest that the inherent capacity first rises to a maximum and then decreases significantly as the heat of reaction increases. This leads to a drop in the inherent capacity at a fixed regenerator temperature, which in turn leads to an overall increase in required equivalent work. In order to maintain constant CO₂ removal rates, the solvent circulation rate increases or overstripping takes place in the regenerator; overstripping is extra heat input necessary to achieve reasonable capacities beyond what is needed for equilibrium considerations. The increased solvent circulation rate leads to greater pumping power requirements, while overstripping moves the regenerator away from performing as an equilibrium process. At a high enough heat of absorption, either of these effects overcome the savings in compressor power, and thus is the reason for an optimum heat of absorption. Figure 3 shows the total equivalent work as a function of regenerator temperature, heat of reaction and water miscibility, while Figure 4 breaks down the total equivalent work into regeneration, compression and pumping work.

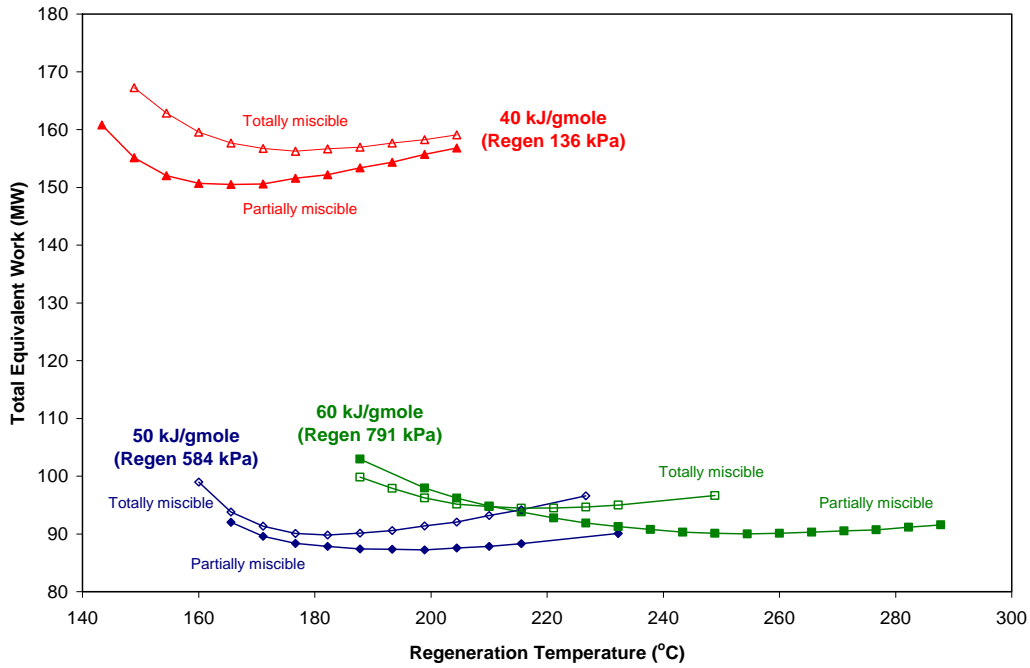


Figure 3. Curves of Lowest Equivalent Work for each ΔH_{rxn} and Water Miscibility.

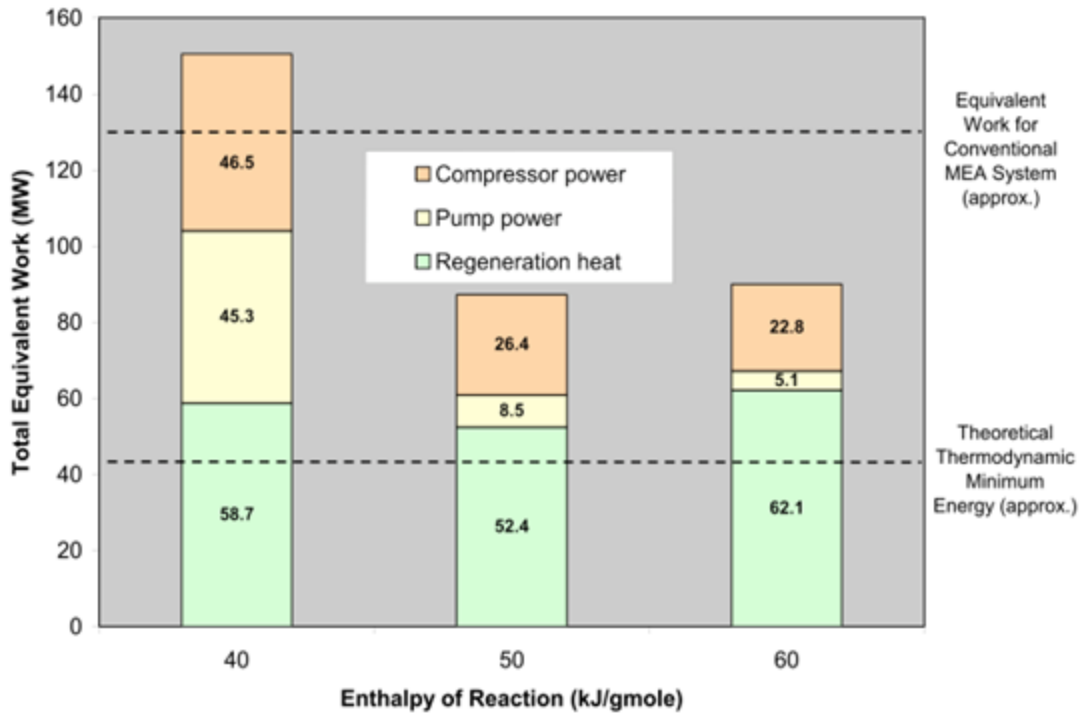


Figure 4. Regeneration Energy and Pumping/Compression Power

CAPITAL AND OPERATING COSTS

Table 2 shows a comparison of the purchased equipment costs for the IL and amine-based absorption-stripping process. The cost of the IL CO₂ capture equipment is slightly lower for the higher regenerator pressure cases (Case 3 and 4) than for the low pressure case (Case 1). This is because the CO₂ capture equipment in the higher pressure cases is generally smaller due to lower liquid circulation rates. Price increases for somewhat higher pressure-rated equipment are minimal since the operating pressure in these cases only varies from ~35 to 689 kPa (5 to 100 psig). Case 2 has the highest circulation rate and resulting purchased equipment cost for the CO₂ capture equipment. Compression train costs, as expected, are proportional to the power requirements for the cases. The cost for dehydration is relatively small and about the same for all of the options in the study so a detailed effort to size and select the dehydration equipment was not necessary. The MEA base line costs are estimated to be the highest because of the large liquid circulation rate and CO₂ compression power requirements compared to the IL cases.

Table 2. Purchased Equipment Costs.

Description	Units	MEA	Ionic Liquids			
		Base Case 110C (230F) 170 kPa (10 psig)	Case 1 149C (300F) 136 kPa (5 psig)	Case 2 166C (330F) 584 kPa (70 psig)	Case 3 199C (390F) 584 kPa (70 psig)	Case 4 204C (400F) 791 kPa (100 psig)
Process Description						
Number of trains		4	4	4	4	4
Total inlet flue gas flow rate	kgmol/hr	84,770	84,770	84,770	84,770	84,770
	MMscfd	1,702	1,702	1,702	1,702	1,702
Total circulating liquid flow rate	m ³ /h	26,126	21,457	31,427	19,232	19,862
	gpm	115,043	94,483	138,383	84,683	87,459
Compression	MW	41	42	24	28	21
	hp	54,536	56,297	32,788	37,042	27,835
Steam pressure needed for reboiler	kPa	308	689	990	1,965	2,247
	psi	45	100	144	285	326
Saturated steam temperature	C	126	164	179	212	218
	F	259	328	355	413	425
Purchased Equipment Costs						
Inlet gas equipment cost (blower, cooler)	\$	6,387,000	6,412,000	6,412,000	6,412,000	6,412,000
CO ₂ capture (i.e., absorption/stripping) equipment costs	\$	117,459,000	92,989,000	123,300,000	91,368,000	92,552,000
Compression train costs	\$	26,131,000	18,623,000	12,409,000	13,089,000	11,019,000
Dehydration costs	\$	1,649,000	1,649,000	1,649,000	1,649,000	1,649,000
Total purchased equipment costs	\$	151,626,000	119,673,000	143,770,000	112,518,000	111,632,000

The installed costs for purchased equipment (everything but compression) were estimated using a factored method similar to that reported in chemical engineering literature³. The installed cost factor for compression was based on vendor recommendations for this type of equipment.

Engineering/home office, project contingency, and process contingency were then added to the total process plant cost to arrive at the total plant cost (TPC). The process plant cost was increased by 10% to account for engineering and home office expenses per the DOE/NETL Systems Analysis Guidelines⁴. A

project contingency of 30% was used since the level of project definition falls roughly in the AACE Estimate Class 3 for budget authorization⁴. A process contingency of 5% was used for all of the cases since the absorption/stripping technology is a commercial process and this same factor was used by EPRI in other CO₂ capture studies^{4,5}. An interest and adjustment factor of 10% of the process plant costs PPC was used to arrive at the total plant investment (TPI). This factor was also similar to other EPRI work in the CO₂ capture area⁵.

Per the DOE/NETL SAG⁴, the total capital requirement (TCR) is the total of the total plant investment prepaid royalties, initial catalyst and chemical inventory, startup costs, spare parts, working capital and land. Table 3 shows how the total capital requirement was derived from the process plant cost.

Table 3. Total Capital Requirement

								MEA		Ionic Liquids			
								Base Case		Case 1	Case 2	Case 3	Case 4
								110C (230F) 170 kPa (10 psig)		149C (300F) 136 kPa (5 psig)	166C (330F) 584 kPa (70 psig)	199C (390F) 584 kPa (70 psig)	204C (400F) 791 kPa (100 psig)
Process Plant Cost (PPC)								329,176,800	258,286,400	302,722,200	239,485,200	236,104,200	
Total Plant Cost (TPC)													
Engineering cost		% of PPC	10	\$	32,918,000	25,829,000	30,273,000	23,949,000	23,611,000				
Process Contingency		% of PPC	5	\$	16,459,000	12,915,000	15,137,000	11,975,000	11,806,000				
Project Contingency		% of PPC	30	\$	98,754,000	77,486,000	90,817,000	71,846,000	70,832,000				
Total Plant Cost (TPC)				\$	477,308,000	374,517,000	438,950,000	347,256,000	342,354,000				
Total Plant Investment (TPI)													
Interest and Inflation Factor		% of PPC	10	\$	32,918,000	25,829,000	30,273,000	23,949,000	23,611,000				
Total Plant Investment				\$	510,226,000	400,346,000	469,223,000	371,205,000	365,965,000				
Total Capital Requirement (TCR)													
Prepaid royalties		% of PPC	0.5	\$	1,646,000	1,292,000	1,514,000	1,198,000	1,181,000				
Initial catalyst and chemical inventory		days inventory	30	\$	18,000	1,000	1,000	1,000	1,000				
Startup Costs													
Component 1		% of TPI	2	\$	10,205,000	8,007,000	9,385,000	7,425,000	7,320,000				
Component 2		days chemicals and operating labor	30	\$	210,000	118,000	121,000	112,000	112,000				
Component 3		days fuel inventory	7.5	\$	0	0	0	0	0				
Spare parts		% of TPC	0.5	\$	2,387,000	1,873,000	2,195,000	1,737,000	1,712,000				
Working Capital													
Component 1		days fuel and consumables	30	\$	204,000	112,000	114,000	105,000	105,000				
Component 2		days byproduct inventory	30	\$	0	0	0	0	0				
Component 3		days direct expenses	30	\$	922,000	726,000	849,000	674,000	664,000				
Land		1960	\$/acre										
		20	acres	\$	39,200	39,200	39,200	39,200	39,200				
Total Capital Requirement				\$	525,857,200	412,514,200	483,441,200	382,496,200	377,099,200				

The major operating and maintenance (O&M) costs for the CO₂ capture and compression process include consumables, maintenance, and plant labor. Byproduct credits, if there are any, are not included. The operating costs are based on a generic site location and should represent a reasonable average of those in various regions of the country. The O&M cost factors, except for operating labor, were obtained from the Systems Analysis Guidelines⁴. The variable O&M costs were specific to the operation

of the CO₂ capture and compression system and depend on the capacity factor (or load factor) of the plant. A capacity factor of 80% was used as recommended in the Systems Analysis Guidelines. The variable O&M components include costs of chemicals consumed (from degradation, entrainment, spills, etc.), utilities, and services used. Solvent losses are not included in the O&M cost estimates for either MEA or IL since data are not available for the final IL solution. In addition, limited experimental data are available to provide upper and lower bounds for the IL degradation losses, but IL degradation losses will need to be researched more thoroughly since the degradation costs could have a large impact on the overall economics of the absorption/stripping process.

The solid waste disposal cost is based on systems typically used for amine units until better information is available for the IL systems. The cost includes such items as activated carbon replacement, assuming that an activated carbon bed will remove some of the compounds formed from the degenerated IL. It is assumed that the carbon beds will be replaced every 3-6 months at an estimated consumption rate of about 0.075 kg C/tonne CO₂ and the cost for the solid waste disposal is \$175/tonne waste⁶. A cooling water system is included in the capital costs and so only makeup water requirements are considered as an operating expense. The O&M costs are shown in Table 4.

Table 4. Summary of Operating and Maintenance Costs

				Ionic Liquids				
				MEA	Case 1	Case 2	Case 3	Case 4
				Base Case 110C (230F) 170 kPa (10 psig)	149C (300F) 136 kPa (5 psig)	166C (330F) 584 kPa (70 psig)	199C (390F) 584 kPa (70 psig)	204C (400F) 791 kPa (100 psig)
Total Operating Costs (TOC)								
Consumables (1 year at capacity factor)								
Chemicals								
		Solvent unit cost	\$/tonne	1,200	5,000	5,000	5,000	5,000
		Solvent usage	kg/h	20.7	0.0	0.0	0.0	0.0
		Solvent cost per year - CO2 capture	\$/year	174,000	1,000	1,000	1,000	1,000
Water								
		Makeup water rate	gpm	5845	3408	3488	3205	3203
		Water unit cost	\$/1000 gal	0.92	0.92	0.92	0.92	0.92
		Water cost per year - CO2 capture	\$/year	2,261,000	1,319,000	1,350,000	1,240,000	1,240,000
		Mercury removal (activated carbon)		0	0	0	0	0
Waste disposal								
			\$/tonne	175	175	175	175	175
			kg waste/ tonne CO2 captured	0.075	0.075	0.075	0.075	0.075
			tonne/year	218	186	188	190	191
			\$/year	39,000	33,000	33,000	34,000	34,000
		Fuel	\$/year	0	0	0	0	0
		Total consumables - CO2 Capture	\$/year	2,474,000	1,353,000	1,384,000	1,275,000	1,275,000
Maintenance Costs								
		Factor	% of TPC	2.2	2.2	2.2	2.2	2.2
		Total maintenance costs	\$/year	10,501,000	8,240,000	9,657,000	7,640,000	7,532,000
Plant Labor								
		Operating labor	1 operator/year	80,000	80,000	80,000	80,000	80,000
		Supervisory and clerical labor						
		Component 1	% of operating labor	30	30	30	30	30
			\$/year	24000	24000	24000	24000	24000
		Component 2	% of maintenance costs	12	12	12	12	12
			\$/year	1,261,000	989,000	1,159,000	917,000	904,000
		Total labor costs	\$/year	1,365,000	1,093,000	1,263,000	1,021,000	1,008,000
		Total Operating Costs	\$/year	14,340,000	10,686,000	12,304,000	9,936,000	9,815,000
Byproduct Credits								
		Unit price for CO2	\$/tonne	0	0	0	0	0
		Total byproduct credits	\$/year	0	0	0	0	0
		Net Operating Cost - CO2 Capture	\$/year	14,340,000	10,686,000	12,304,000	9,936,000	9,815,000

The energy requirements to operate the main facility and the CO₂ capture unit are withdrawn from the main power plant output either through electricity or steam. This decreases the net electrical output of the plant. Power requirements of electric motors (i.e., inlet blower, pumps, and electric compressor drivers for the IL cases) translate directly to electrical derating (a decrease in MW_e). Energy requirements that are supplied using steam, such as the heat requirements for the reboilers and steam compressor drivers (MEA base case only), must be converted into electrical derating by calculating the amount of electrical generating capacity that the steam would have supplied to the main power facility had the steam not been diverted to the CO₂ capture system.

The steam taken from the turbine will be de-superheated with boiler feed water so somewhat less steam will be drawn off from the plant than actually required in the CO₂ capture process. This is taken into account in the reboiler derating calculations. The amount of steam needed for the reboiler in the IL cases was estimated to be 41 to 48% less than the amount of steam needed for a MEA system. Table 5 summarizes the electrical derating that results from the energy requirements for the four IL cases and the MEA base case as well as the net generating capacity for the 500 MW_e gross plant with and without CO₂ capture.

The CO₂ capture derating for the IL cases (100 to 110 MW_e) was estimated to be lower than the MEA base case (132 MW_e) because of the large steam demand in the MEA reboiler (2-2.5 times more than for the IL cases). The IL-based process decreases the overall derating of CO₂ capture by 13 to 18%.

Table 5. Derating Results

Description	Units	MEA	Ionic Liquids			
		Base Case	Case 1	Case 2	Case 3	Case 4
		110C (230F) 170 kPa (10 psig)	149C (300F) 136 kPa (5 psig)	166C (330F) 584 kPa (70 psig)	199C (390F) 584 kPa (70 psig)	204C (400F) 791 kPa (100 psig)
Base Plant Power	MWe	500	500	500	500	500
Main facility derating	MWe	29.1	29.1	29.1	29.1	29.1
Inlet blower work (direct electricity usage)	MWe	8.4	8.4	8.4	8.4	8.4
Pump work (direct electricity usage)	MWe	14.7	11.5	15.2	10.2	12.2
Reboiler heat (steam derating)	MWe	118	45	51	56	59
Compressor work (supplemental electric driver)	MWe	0	42	24	28	21
CO ₂ pump work	MWe	1.9	3.2	3.4	3.2	3.2
Excess energy from turbine	MWe	-10.8	0	0	0	0
CO ₂ capture derating	MWe	132.0	110.0	102.8	105.8	103.3
Total derating	MWe	161.0	139.1	131.8	134.9	132.3
Plant Net Electrical Capacity - without capture	MWe	471	471	471	471	471
Plant Net Electrical Capacity - with capture	MWe	339.0	360.9	368.2	365.1	367.7

ECONOMIC ANALYSIS AND CONCLUSIONS

Table 6 shows the cost of electricity with and without CO₂ capture for the four ionic liquid cases and the MEA base case. The base plant cost of electricity is assumed to be 5 ¢/kWh. The cost of electricity is the highest for the MEA base case (10.6 ¢/kWh). The increase in cost of electricity for the base case is estimated at 113%. The cost of electricity for the four IL cases ranges from 8.8 to 9.5 ¢/kWh with an increase in cost of electricity of 79 to 90%. CO₂ capture with ionic liquids was estimated to be lower by 11 to 17% compared to the MEA base case.

Table 6. Cost of Electricity (excluding costs for solvent losses)

Description	Units	MEA	Ionic Liquids			
		Base Case	Case 1	Case 2	Case 3	Case 4
		110C (230F) 170 kPa (10 psig)	149C (300F) 136 kPa (5 psig)	166C (330F) 584 kPa (70 psig)	199C (390F) 584 kPa (70 psig)	204C (400F) 791 kPa (100 psig)
Gross generating capacity	MWe	500	500	500	500	500
Net generating capacity without CO ₂ capture	MWe	471	471	471	471	471
Net generating capacity with CO ₂ capture	MWe	339	361	368	365	368
Base plant cost of electricity	c/kWh	5.0	5.0	5.0	5.0	5.0
Annual base plant costs	MM\$/yr	165.0	165.0	165.0	165.0	165.0
Total annual CO ₂ capture costs	MM\$/yr	87.8	68.4	80.0	63.5	62.6
Total annual costs with CO ₂ capture	MM\$/yr	252.9	233.5	245.0	228.6	227.6
Total COE	c/kWh	10.6	9.2	9.5	8.9	8.8
Increase in COE	%	113	85	90	79	79

The cost of electricity and the cost of CO₂ avoided are provided in Table 7 for three processes: MEA absorption-stripping, IL absorption-stripping and IL thermal swing adsorption. The data in Table 7 clearly illustrate that IL adsorption (TSA) is economically unattractive when compared to absorption-stripping alternatives. The excessive electricity and CO₂ capture costs stem from the large plant derating, which is related to the unrecoverable heat losses from the heating the subsequent cooling of the silica gel pellets during each adsorption cycle.

Table 7. IL COE/Cost of CO₂ Avoided Summary

	Units	MEA Base Case	IL Absorption-Stripping	IL TSA Base	IL TSA Optimized
Net generating capacity with CO ₂ capture	MW _e	339	368	117	188
Total annual costs with CO ₂ capture	MM\$/yr	253.6	227.6	354.4	318.0
Total COE with CO ₂ capture	c/kWh	10.7	8.8	43.4	24.1
Increase in COE	%	113	79	768	383
Cost of CO ₂ avoided	\$/tonne	67.20	44.84	657.06	260.40

The economic comparison shows that ionic liquid solutions could potentially provide significant savings to the utility industry over standard MEA technology for CO₂ capture. However, much research is still being done to develop the optimal ionic liquid solution for CO₂ capture and the results in this study should be considered preliminary in nature. Several key assumptions were used in the economic analyses that could impact the overall results. The most important of these assumptions is the degradation rate of the ionic liquid solution. As mentioned earlier, degradation was not included in the

economic comparisons presented in this report. This is because preliminary research has identified a very wide range of possible degradation rates depending on the choice of ionic liquid. The degradation basis would need to be 2.4 kg/tonne CO₂ for the IL solution to remain at the same cost of electricity as the MEA solution for Case 2.

Another important factor in the economic analysis is the material of construction used for the equipment in the CO₂ capture process. Corrosion data are not yet available for the ionic liquids and so it was assumed that the same material of construction could be used as with a MEA unit in CO₂ capture service. If the corrosion rates are significantly better or worse than different materials could be used that will impact the capital cost of the process.

The kinetics for CO₂ absorption will affect the height of the packing required in the absorber and stripper towers. For now, it was assumed that CO₂ will be absorbed in the ionic liquid at the same rate as MEA and, thus, similar packing heights were used. Solution upkeep (i.e., filters and reclamation) is assumed to be the same as with MEA, too.

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