INTEGRATING MEA REGENERATION WITH CO_2 COMPRESSION AND PEAKING TO REDUCE CO_2 CAPTURE COSTS

Final Report of

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EXECUTIVE SUMMARY

Capturing CO_2 from coal-fired power plants is a necessary component of any large-scale effort to reduce anthropogenic CO_2 emissions. Conventional absorption/stripping with monoethanolamine (MEA) or similar solvents is the most likely current process for capturing CO_2 from the flue gas at these facilities. However, one of the largest problems with MEA absorption/stripping is that conventional process configurations have energy requirements that result in large reductions in the net power plant output. Several alternative process configurations for reducing these parasitic energy requirements were investigated in this research with the assistance of the Platte River Power Authority, based on recovering energy from the CO_2 compression train and using that energy in the MEA regeneration step. In addition, the feasibility of selective operation of the amine system at a higher CO_2 removal efficiency during non-peak electricity demand periods was also evaluated.

Four process configurations were evaluated: A generic base case MEA system with no compression heat recovery, CO_2 vapor recompression heat recovery, and multipressure stripping with and without vapor recompression heat recovery. These configurations were simulated using a rigorous rate-based model, and the results were used to prepare capital and operating cost estimates. CO_2 capture economics are presented, and the cost of CO_2 capture (cost per tonne avoided) is compared among the base case and the alternative process configurations.

Cost savings per tonne of CO_2 avoided ranged from 4.3 to 9.8 percent. Energy savings of the improved configurations (8 – 10 %, freeing up 13 to 17 MW of power for sale to the grid based on 500 MW unit) clearly outweighed the modest increases in capital cost to implement them; it is therefore likely that one of these improved configurations would be used whenever MEA-based (or similar) scrubbing technologies are implemented. In fact, the payback on capital for the most promising heat integration configurations (Cases 3 and 4) is only six months to one year (based on $0.06/kWh).

Another significant result is that the reboiler steam requirement could be reduced by up to 39% with the advanced process configurations. Selective operation of the amine system was found to be economic only if the value of peak electricity was in excess of approximately $230/MWh (from the assumed $130/MWh to buy power from a supplemental natural gas peak turbine) and, therefore, is not considered to be a reasonable option for minimizing CO_2 capture costs.

These results indicate an improvement to commercial MEA-based technologies, which helps to incrementally meet DOE's Sequestration Program targets when coupled with other process improvements. For example, DOE's target goal of $20/tonne of CO_2 could potentially be achieved by combining use of the heat integration configurations evaluated in this study and other advanced amine solvents (instead of conventional MEA) that have been developed to further reduce the reboiler duty steam requirements. It is expected that the advanced amines could add another 15% savings in cost of CO_2 captured. In addition, advanced aqueous-based solvent approaches already exist and may be commercialized more quickly than other approaches.

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1.0 INTRODUCTION

This report documents the methodology and results of Trimeric Corporation's Small Business Innovative Research (SBIR) Phase I project, "Integrating MEA Regeneration with CO_2 Compression and Peaking to Reduce CO_2 Capture Costs" (DOE Grant No. DE-FG02-04ER84111). This section provides background information on the issues that are driving this type of research, a discussion of the research goals and objectives, the project participants, and an overview of the remainder of the document.

1.1 Background

Ratification or approval of the Kyoto Protocol by 141 nations (most recently, Russia's ratification brought the treaty into effect) demonstrates the concern of the international community about how human activity could potentially be contributing to global warming. Climate change science suggests that higher atmospheric concentrations of greenhouse gases (carbon dioxide [CO_2], methane, etc.) have the potential to increase heat retention in atmosphere, potentially resulting in a wide range of effects. Of the anthropogenic greenhouse gases, CO_2 is the primary concern; in 2001, 82.1 percent of total U.S. greenhouse gas emissions consisted of CO_2 from the combustion of fossil fuels (DOE/EIA, 2002). Since the consequences of changes in global climate are potentially very significant, there is strong interest in reducing the amount of CO_2 emitted to the atmosphere by human activity.

To address global warming concerns, President Bush has committed the United States to pursuing a range of strategies. These initiatives were summarized in February 2002 during President Bush's announcement of the Global Climate Change Initiative (GCCI), which has an overall goal of reducing U.S. greenhouse gas intensity by 18% over 10 years. Because electric power production contributes about 40% of U.S. CO_2 emissions (DOE/EPA, 1999), any effort to reduce greenhouse gas intensity virtually must address this sector.

Recognizing that limiting CO_2 emissions from electric power production must be an essential element of any climate change strategy, President Bush and Secretary of Energy Abraham announced the FutureGen initiative in February 2003. FutureGen is aimed at creating a near-zero emissions coal-fired power plant that integrates hydrogen and electricity production along with carbon capture and sequestration. The initial target for FutureGen was CO_2 removal of 90%, with the goal of approaching nearly 100% capture as technology progresses over time. At about the same time (March 2003), DOE presented its technology roadmap and program plan for implementing the President's GCCI; the DOE plan included a goal to "create systems that capture at least 90% of emissions and result in less than a 10% increase in the cost of energy services" (DOE NETL, 2003). Recent DOE NETL presentations indicated that no more than a 20% increase in COE is targeted for post-combustion capture technologies.

In addition to their large amounts of CO_2 emissions, power plants are a relatively attractive target for CO_2 capture and sequestration because they are relatively few in number and emit relatively large amounts of CO_2 from a single location. These characteristics suggest that capturing and sequestering CO_2 from power plant flue gas should achieve economies of scale and be much more cost-effective than performing the same on other smaller, more widely distributed CO_2 sources (e.g., CO_2 emissions from automobiles) and non-utility point sources.

Of the currently available technologies to capture CO_2 from power plant flue gas, amine-based scrubbing using monoethanolamine (MEA) has been determined to be among the most likely near-term options. MEA scrubbing has been determined to be the least expensive of the near-term options in terms of cost per unit of CO_2 captured (IEA, 1994). While other less expensive CO_2 capture technologies may be developed in the future, some of them may be years away from commercial availability (perhaps outside the 10 year window in the GCCI goal). By contrast, MEA scrubbing is used in many non-power applications today, and MEA scrubbing represents a technology option that can be applied to full-scale plants within the next few years.

Simple MEA absorption/stripping processes have been applied on a small-scale to scrub CO_2 from flue gas at several coal-fired and natural gas turbine power plants in the US (Chapel, 1999; Sander and Mariz, 1998) and from engine exhaust (Hopson, 1995). However, there are no large, full-scale, commercial implementations of the technology.

There are several significant challenges with using MEA scrubbing on flue gas. Residual oxygen, SO_2, and other species will cause chemical degradation of the MEA. The MEA liquid solution can be corrosive to process equipment. Finally, and perhaps most significantly, the capital and energy costs to implement MEA scrubbing on power plant flue gas are high.

In terms of energy costs, this study showed that the energy consumption of a simple MEA absorption and stripping process along with CO_2 compression (e.g., for injection) at a power plant may be about 38% of the total power plant energy requirement. Clearly, reducing these parasitic energy requirements is crucial to early application of MEA systems for full-scale CO_2 capture projects.

1.2 Research Objectives

In previous work, research at the University of Texas defined the actual energy, ideal/theoretical energy, and lost work involved with MEA absorption and stripping approaches (Rochelle, 2003). The result was that more than half of the energy required by a standard MEA and CO_2 compression approach was the result of lost work; losses in the MEA stripper were the largest (~70% of total). Several innovative processing approaches were proposed to reduce this lost work (by 5 to 20%); in general, these approaches involved integrating the need for heat in the MEA stripper with the needs of the CO_2 compression train and reducing temperature approaches in the lean/rich exchanger. This project sought to build on that previous research by conducting an engineering and economic analysis of those innovative processing approaches to determine if significant cost savings could be achieved.

The overall objective of this research was to identify additional ways to reduce costs as well as to determine the optimal approach for implementing these energy saving ideas at acceptable capital costs. The specific technical objectives of this project were to:

- Develop process designs for approximately three innovative MEA stripper configurations to reduce parasitic energy requirements of CO_2 capture with MEA;
- Develop and evaluate other novel processing schemes that are discovered as a result of process design, engineering, or integration planning;
- Evaluate equipment options and select equipment with the best combination of operability and economics to implement the process designs; and
- Determine how to best integrate the MEA process and CO_2 compression into a coal-fired utility so as to accomplish 90% CO_2 removal at least cost.

1.3 Project Participants

Trimeric Corporation served as the prime contractor for this project. Dr. Gary Rochelle of the University of Texas and his research group performed the process simulations and provided general technical insight and guidance. Platte River Power Authority (PRPA), a not-for-profit electricity generator that provides power to four cities in northern Colorado, provided input on coal-fired power plant operations and integration of the CO_2 capture system into an existing plant.

1.4 Report Organization

The remainder of this document presents the research performed under this project and is organized as follows:

- Section 2: Conceptual Approach describes the overall design basis and options considered as part of this project;
- Section 3: Process Simulation and Design provides a description of the process modeling and results, including heat and material balances;

- Section 4: Equipment Sizing and Selection discusses how the results of the process simulation were used in selecting equipment and presents the equipment details for each case that was evaluated;
- Section 5: Capital and Operating Costs summarizes the cost of the equipment and operations for the various cases;
- Section 6: Economic Analysis and Results shows a comparison of the costs of the different cases as well as a scenario for not operating the CO_2 recovery system during peaking periods; and
- Section 7: Summary and Conclusions presents the findings of the research.

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2.0 CONCEPTUAL APPROACH

This section discusses the conceptual approach that was used on the project. A discussion of the base plant's design basis, the various innovative processing approaches evaluated as part of this research, and the general engineering analysis approach is presented in this section.

2.1 Process Simulation Design Basis

The base case (also referred to as Case 1) for the monoethanolamine (MEA) process simulation is defined as shown in Table 2-1. The conditions shown in Table 2-1 are for the flue gas exiting the coal-fired power plant and entering the MEA unit. The composition, conditions, and flow rates for the flue gas as well as MEA base case design for the absorber and stripper shown were derived from the previous modeling research performed under guidance from Dr. Gary Rochelle (Freguia, 2002). Key assumptions about the flue gas and MEA system design basis are as follows:

- The composition of the flue gas is based on a conventional pulverized coal (PC) boiler.
- A wet flue gas desulfurization (FGD) scrubber is applied to the flue gas from the coal-fired power plant to achieve both SO_2 removal (to prevent interference with the MEA) and cooling of the inlet gas stream to the CO_2 capture system.
- No interferences from NO_x or other pollutants are expected.
- The MEA system described in Table 2-1 is a typical MEA design, as shown in Figure 2-1.
- CO_2 stream complies with the well class type that will accept the CO_2 for geological sequestration.
- CO_2 quality meets pipeline lifetime expectancy according to industry standards.

Table 2-1. CO_2 Capture Design Basis for Base Case Process Simulation

Flue Gas					
	Composition (mol%)				
	CO2	12.33			
	H2O	9.41			
	N2	73.47			
	O2	4.77			
	Water saturation temperature	47	C	116.6	F
	Absorber inlet temperature	55	C	131	F
	Absorber inlet pressure	111.325	kPa	16.15	psia
	Mole flow (after saturation)	0.0794	kmol/m2-s	0.0162	lbmol/ft2-s
Solvent					
	Unloaded composition				
	MEA (30 wt%)	11.23	mol%		
	H2O	88.77	mol%		
	Lean loading (mol CO2/mol MEA)	adjusted to minimize energy requirement			
	Lean solvent temperature	40	C	104	F
	CO2 removal	90	%		
	(Solvent rate is calculated to get specified removal)				
Absorber					
	Packing height	15	m	49.2	ft
	Diameter	7	m	23.0	ft
	Pressure drop	10	kPa	1.5	psia
	CO2+MEA kinetics	From Dang (2001)			
	Packing type	Cascade mini rings #2			
Cross Exchanger					
	Temperature approach, hot end	10	C		
Stripper					
	Packing height	10	m	32.8	ft
	Diameter (based on 80% of flooding)	4.5	m	14.8	ft
	Bottom pressure	172.12	kPa	25.0	psia
	Pressure drop	10	kPa	1.5	psia
	Reboiler	Equilibrium			
	Condenser (equilibrium partial condenser)	50	C	122	F
	Rich solvent feed location (from top)	0.5	m	1.6	ft
	Water reflux location	At top			
	Packing type	Cascade mini rings #2			
	Reactions	All at equilibrium			

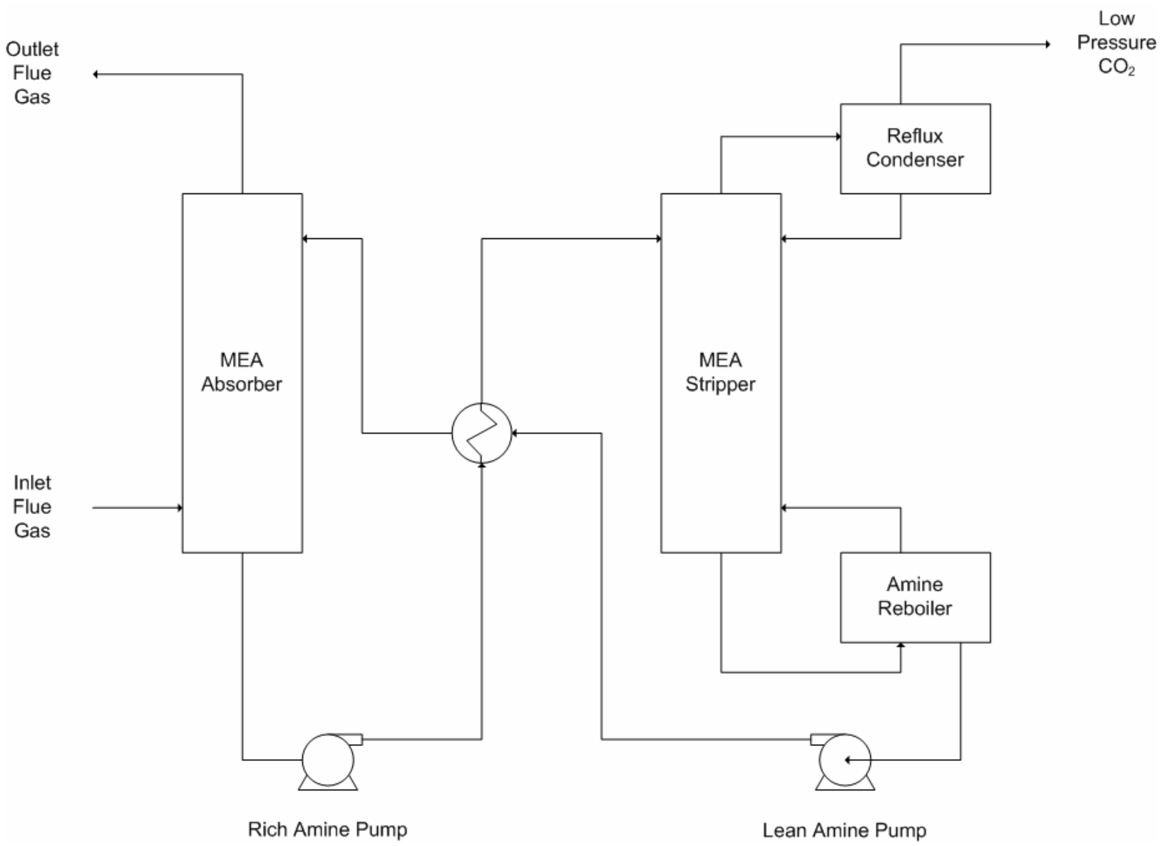

Figure 2-1. Typical MEA Unit

2.2 Process Configurations

Figure 2-2 presents the base case design that was used for this project. Previous research suggested that more than half of the energy required by a standard MEA and CO_2 compression approach was the result of lost work; losses in the MEA stripper were the largest (Rochelle, 2003). Several innovative processing approaches were proposed to reduce this lost work by up to 20%; in general, these approaches involved integrating the need for heat in the MEA stripper with the needs of the CO_2 compression train and reducing temperature approaches in the lean/rich exchanger. Subsequent analysis by Dr.

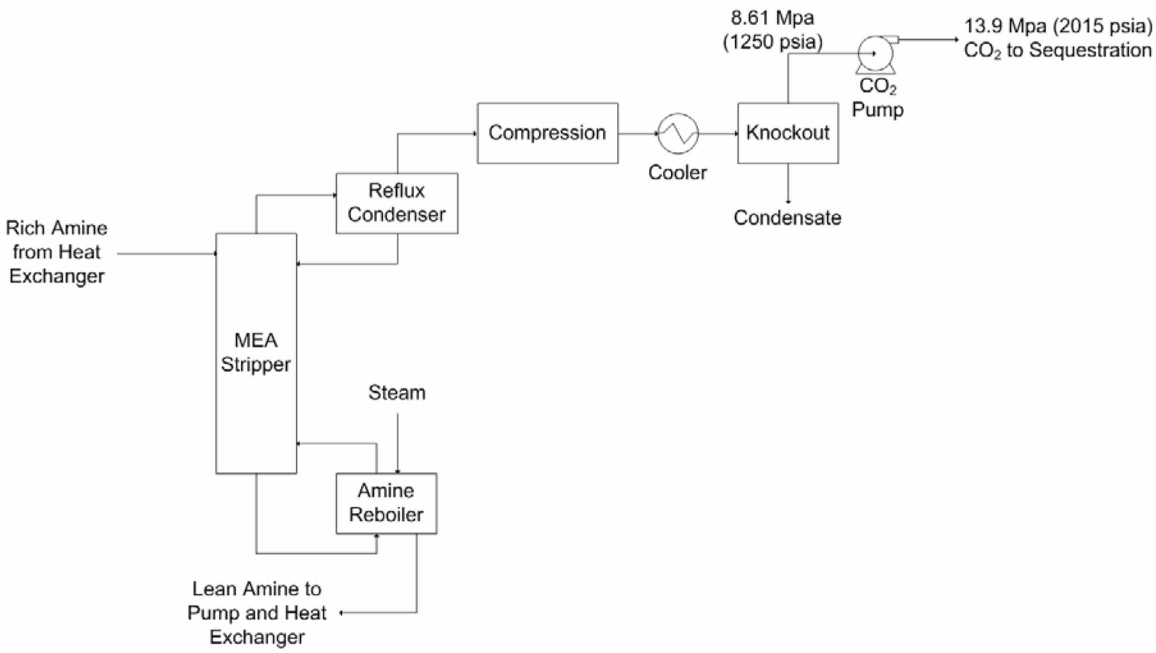

Figure 2-2. Case 1 and 1b: CO_2 Compression off MEA Stripper

Rochelle suggested that reducing the temperature approach in the lean/rich exchanger may result in a pinch point on the cold end of the exchanger that limits the minimum approach temperature. As a result, this project focused on integrating the MEA stripper with the CO_2 compression. The cases evaluated for this project are described below.

2.2.1 Cases 1 and 1b: Base Case MEA Unit with CO_2 Compression at 90% and 95% Recovery

Cases 1 and 1b are shown in Figure 2-2 above. These cases include the basic MEA unit followed by compression of the CO_2 to 8.6 MPa (1250 psia), then cooling the CO_2 with water and pumping the dense phase CO_2 up to 13.9 MPa (2015 psia) for transport. Case 1 is based on 90% CO_2 removal, while Case 1b is the same as 1 except it has 95% CO_2 removal.

The purpose of Case 1b is to allow evaluation of selective operation of the MEA unit. The strategy of selective operation of the amine system (with its large power consumption) involves operating at higher than 90% reduction (e.g., 95+% CO_2 capture) during periods when power demand is lower, and then shutting down the amine system and maintaining it on hot standby for some fraction of time (e.g., 5%) during peak demand periods when the power demand is highest. As a result, this enables the plant to achieve an overall CO_2 recovery of ~90% on an annualized basis, while minimizing the addition of generation capacity that is needed if CO_2 is controlled. An additional possible benefit of selective operation is that the peaking periods typically correspond with summertime operations, when the cooling requirements to achieve the desired lean amine temperature are greatest; by not operating under the most extreme conditions, the amine unit design could potentially be slightly scaled back, with the potential for some capital cost savings achieved.

2.2.2 Case 2: Heat Recovery

Figure 2-3 illustrates Case 2. This case is similar to Case 1, but with two significant differences in the process. Heat recovery is achieved by eliminating the reflux condenser, compressing the entire stripper overheads stream up to 8.6 MPa (1250 psia) with multistage compression, and using the hot compressor discharge stream from each stage as a heat source for the amine reboiler. After passing through the amine reboiler and being cooled, condensate (water) is recovered from the CO_2 stream and recycled to the process, and the dense CO_2 is pumped up to pipeline pressure.

2.2.3 Case 3: Multipressure Stripping with Heat Recovery

Figure 2-4 outlines Case 3. This case includes the heat recovery of Case 2 but with vapor recompression added into the stripper. Essentially, the stripper is modified to integrate the first two stages of compression into the stripper. All of the vapors from the stripper are compressed and reinjected at the next higher pressure as the vapor progresses up the column. The bottom of the stripper operates at approximately 202.6 kPa (29.4 psia), the middle section operates at 283.7 kPa (41.2 psia), and the top of the stripper operates at 405.3 kPa (58.8 psia).

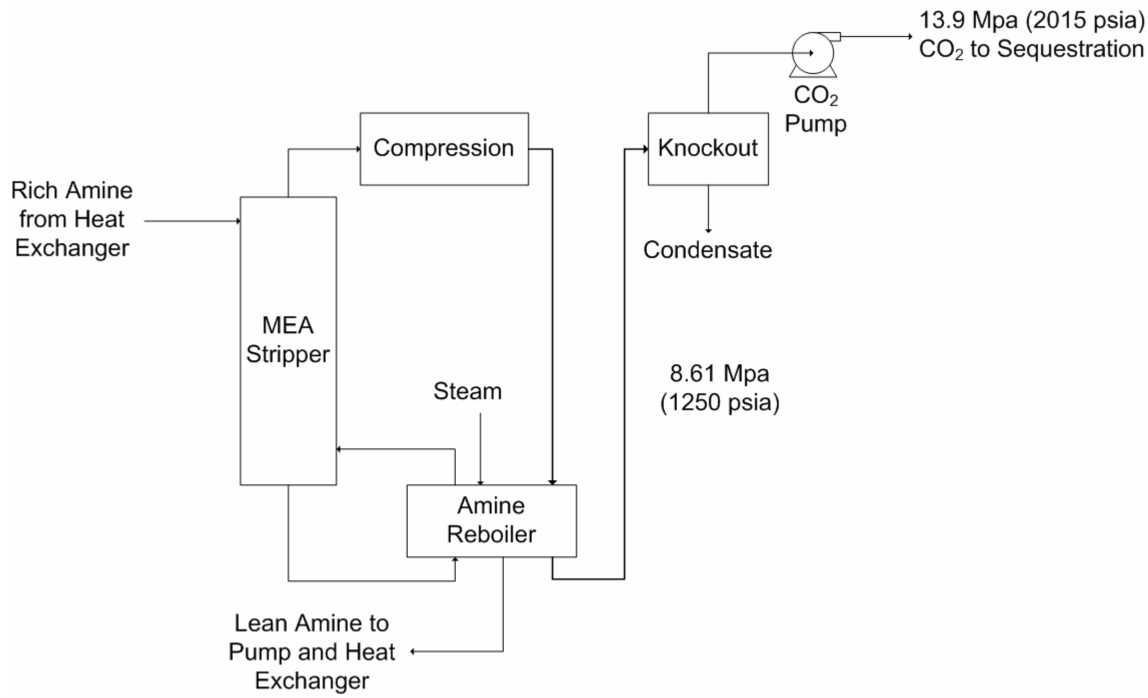

Figure 2-3. Case 2: CO_2 Compression with Heat Recovery

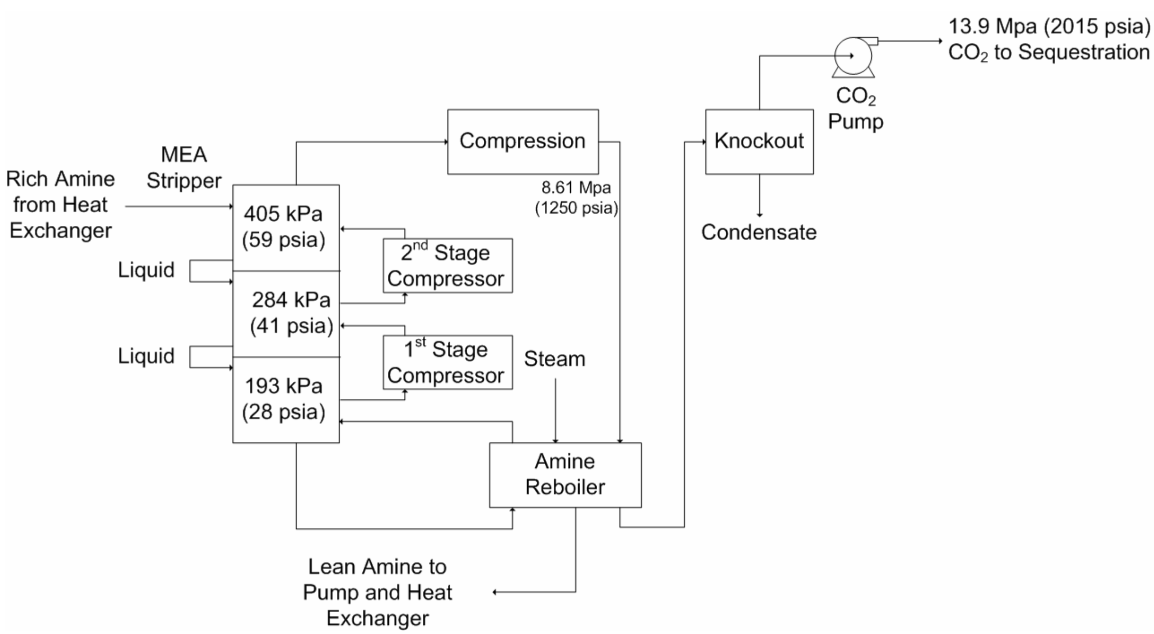

Figure 2-4. Case 3: Multipressure Stripping with Heat Recovery

2.2.4 Case 4: Multipressure Stripping without Heat Recovery

Figure 2-5 provides a block flow diagram of Case 4. Case 4 is very similar to Case 1 (both the reflux condenser and compressor discharge are cooled with water and no heat recovery from the later stages of compression is integrated into the process), but operates with a multipressure stripper as described in Case 3.

2.3 Engineering and Economic Analysis Approach

This section gives a brief overview of the process simulation approach, the scaling of the simulation results, equipment sizing, and economic analysis performed during the project. A more detailed discussion of these areas follows in subsequent sections of this report.

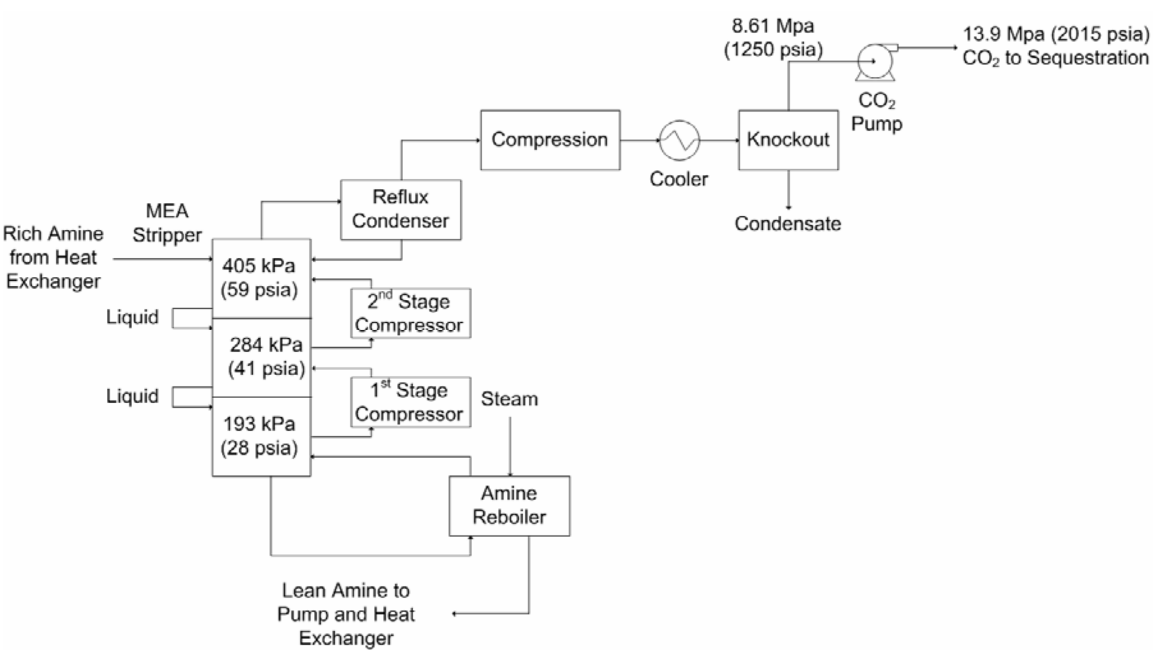

Figure 2-5. Case 4: Multipressure Stripper without Heat Recovery

2.3.1 Process Simulation

The University of Texas at Austin conducted process simulations using the design basis described in Section 2.1 for the four cases listed in Section 2.2. The simulations used Aspen Plus with RateFrac for both the absorber and stripper. The absorber was modeled with kinetic reactions, while the stripper modeling used all equilibrium reactions. The NRTL model for electrolyte solutions was used both for calculating equilibrium in the stripper and for calculating activities for the kinetic modeling in the absorber.

2.3.2 Scaling of Simulation Results

Once the process simulations were completed, the engineering evaluation could be conducted. Before mass and energy balances could be performed for the CO_2 removal system, the simulation results needed to be scaled to a size that would be representative of full-scale coal-fired power plants. Scaling the simulation involved selecting values for gross plant capacity, gross plant heat rate, coal carbon content, and coal heating value.

As noted in Section 2.1, it was assumed that a wet FGD system (i.e., limestone scrubbing) was applied to the flue gas stream prior to the stream entering the MEA unit. According to the 2001 EIA-767 database, the average capacity for coal-fired utility plants with limestone FGD systems was 497 megawatts (MW). Therefore, 500 MW was selected as the gross plant capacity.

For the remaining values, a gross heat rate of 9,674 Btu/kWh was chosen based on recent EPRI data (EPRI 2000). The coal composition and fuel heating value were based on guidelines from DOE for Illinois #6 coal (DOE, 2004).

Having chosen these values, the CO_2 flow rate was calculated for a 500 MW facility. This flow was used to scale the results of the process simulations up to the 500 MW size.

2.3.3 Equipment Sizing

Once the process simulation results were scaled to a 500 MW plant, heat and material balances were calculated, and equipment specifications and sizing were performed. Section 4 and 5 of this report provide an in-depth discussion of the methodologies used.

2.3.4 Economic Analysis

Sections 5 and 6 of this report provide greater detail on the development of capital and operating costs and the economic comparison of the different cases. However, in developing these costs, certain assumptions were made about the site and type of utility operations involved. These assumptions included the following:

- The coal-fired power plant is assumed to be a base-load power plant that is central to the utility's electrical generating system rather than an intermediate (or "swing") load unit or a peaking unit. Based on this, an 85% capacity factor was used for the economic analyses. (A sensitivity analysis to the capacity factor is discussed in Section 6).
- The CO_2 capture system installation is assumed to be a retrofit to an existing power plant, since this would describe the bulk of the systems that may be installed.
- The CO_2 removed by the MEA unit is compressed to a pipeline pressure of 13.9 MPa (2015 psia) for transport and injection at an off-site location.
- Dehydration or other treatment of the CO_2 is not included for the purposes of this comparison for two reasons. First, the comparisons in this project are limited to different configurations of the MEA system; any processing after the MEA unit would be the same for all cases evaluated in this project.

Secondly, dehydration of CO_2 may or may not be required depending on the specific sequestration approach.

Typical comparison costs, such as the cost per ton CO_2 avoided and the effect of CO_2 removal systems on the costs of electricity, were developed and are presented in Section 6.

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3.0 PROCESS SIMULATION AND DESIGN

This section describes the results of the process simulation and design task. The goal of the process simulation work was to generate heat and material balances for the multiple MEA stripper configurations investigated in this study. The heat and material balances were then used as a basis for the subsequent equipment sizing, selection, and economic evaluation tasks.

3.1 Process Simulation Approach

The primary process simulations were developed using Aspen Technology Inc.'s Aspen Plus, version 12.1, with the RateFrac module for modeling the absorber and the stripper. Some of the ancillary processes, e.g., the steam desuperheating, cooling water system, and the CO_2 compression trains, were modeled separately using WinSim's Design II, version 9.17. All of the process calculations were based on steady-state conditions at the full design capacity of the unit for each case. The following subsections describe the scope of the simulations, the thermodynamic and physical property specifications, and the major process specifications used to build the simulations.

3.1.1 Simulation Scope

The scope of the simulations was limited to the CO_2 capture and compression equipment. The scope excluded simulations of the utility power generation system and non-CO_2 pollution control equipment such as flue-gas desulfurization (FGD) units, electrostatic precipitators (ESPs), and selective catalytic reduction (SCR) units. The feed stream for the simulation was the flue gas stream just prior to the CO_2 absorber and downstream of any flue gas blowers and pollution control equipment. The simulation included the entire MEA system, which consists of an absorber, regenerator, associated process heat exchangers and pumps, and CO_2 compression train including all interstage coolers and separators. CO_2 dehydration equipment was not included in the process simulation because 1) the costs would be the same among the various configurations under comparison, and 2) dehydration may not be required in cases where the CO_2 capture equipment is located near a subsurface sequestration site. The simulation terminated with a CO_2 product delivered to the battery limits at 13.9 MPa (2000 psig) and approximately 52°C (125˚F).

3.1.2 Thermodynamic and Physical/Chemical Properties Specifications

The RateFrac model used in this study was originally developed by Freguia (2002) with minor modifications. The model assumes instantaneous reactions in the stripper and finite reaction rates in the absorber, and includes the effects of liquid-phase and gas-phase diffusion resistances.

The model represents vapor-liquid equilibrium and solution speciation with the NRTL electrolyte model regressed on the MEA data of Jou and Mather (1995). The reactions included in the absorber RateFrac model are shown in the following seven equations:

$$H_2O + MEA^{+} \longleftrightarrow H_3O^{+} + MEA \qquad (1)$$

$$2\ H_2O \longleftrightarrow H_3O^{+} + OH^{-} \qquad (2)$$

$$H_2O + HCO_3^{-} \longleftrightarrow H_3O^{+} + CO_3^{2-} \qquad (3)$$

$$CO_2 + OH^{-} \longrightarrow HCO_3^{-} \qquad (4)$$

$$HCO_3^{-} \longrightarrow CO_2 + OH^{-} \qquad (5)$$

$$H_2O + CO_2 + MEA \longrightarrow MEACO_2^{-} + H_3O^{+} \qquad (6)$$

$$MEACOO^{-} + H_3O^{+} \longrightarrow H_2O + CO_2 + MEA \qquad (7)$$

Equations one through three are equilibrium equations; equations four through seven are kinetic equations. Kinetic rate coefficients were based on data from Dang and Rochelle (1991). The reactions included in the stripper RateFrac model are shown in the following five equations:

$H_2O + MEA^+$	$\longleftrightarrow$	$H_3O^+ + MEA$	(1)
$2\ H_2O$	$\longleftrightarrow$	$H_3O^+ + OH^-$	(2)
$H_2O + HCO_3^-$	$\longleftrightarrow$	$H_3O^+ + CO_3^{2-}$	(3)
$2\ H_2O + CO_2$	$\longleftrightarrow$	$H_3O^+ + HCO_3^-$	(8)
$H_2O + MEACOO^-$	$\longleftrightarrow$	$MEA + HCO_3^-$	(9)

All five equations are equilibrium equations, which corresponds to instantaneous reactions in the stripper. Equations one through three are common to both the absorber and the stripper.

The physical and thermodynamic property methods used are summarized below:

- Vapor heat capacities – Vapor heat capacities were based on the DIPPR correlation for non-electrolyte species and on a polynomial form for electrolyte species.
- Heats of vaporization- Heats of vaporization were based on the DIPPR correlation for non-electrolytes and on the Watson correlation for electrolytes.
- Liquid densities – Liquid densities were based on the DIPPR correlation.
- Vapor and supercritical fluid densities – Soave-Redlich-Kwong (SRK) equation of state.
- Diffusivities – Diffusivities used the Chapman-Enskog-Wilke-Lee model for mixtures.
- Thermal conductivities – Thermal conductivities used DIPPR correlations.
- Viscosities – Viscosities were based on the DIPPR model for non-electrolytes and on the Andrade correlation with the Jones-Dole correction for electrolyte species.
- Surface tension – Surface tensions were based on the DIPPR correlation.
- Solubility of supercritical components - CO_2, N_2, and O_2 were modeled using a Henry's Law correlation.

3.1.3 Key Process Simulation Specifications

Table 3.1 presents the key process simulation inputs for each of the six cases. These inputs include the flow rates and compositions for feed streams as well as required conditions for each of the unit operations. The differences between the cases are values for lean amine circulation rate, absorber diameter, rich amine pump discharge pressure, stripper diameter, stripper reboiler duty, reflux condenser pressure, and number of compression stages. Absorbers and strippers for all cases use the same packing: cascade mini rings # 2.

Table 3-1. Summary of Process Simulation Inputs

Description	Unit	Value					
		Case 1	Case 2	Case 3	Case 4	Case 1b	Case 2b
Inlet Flue Gas							
Flow rate	kgmol/h	84,827	=	=	=	=	=
Temperature	C	55	=	=	=	=	=
Pressure	kPa	111.2	=	=	=	=	=
Composition							
H_2O	molefrac	0.0941	=	=	=	=	=
CO_2	molefrac	0.1233	=	=	=	=	=
N_2	molefrac	0.7349	=	=	=	=	=
O_2	molefrac	0.0477	=	=	=	=	=
Lean Amine (Absorber Feed)							
Circulation rate	L/s	2798	2794	2854	2857	3347	3339
Temperature	C	40	=	=	=	=	=
MEA Makeup							
Flow rate	kgmol/h	0.067	=	=	=	=	=
Temperature	C	37.8	=	=	=	=	=
Absorber							
CO_2 removal	%	90	90	90	90	95	95
Packing type	-	CMR (Cascade Mini Rings)	=	=	=	=	=
Packing arrangement	-	Random	=	=	=	=	=
Packing material	-	stainless steel	=	=	=	=	=
Packing size	cm	3.81	=	=	=	=	=
Packing specific surface area	m2/m3	144.0	=	=	=	=	=
Packing factor	1/m	85.3	=	=	=	=	=
Surface tension of packing	dyne/cm	75.0	=	=	=	=	=
Void fraction	-	0.971	=	=	=	=	=
Packing height	m	15.0	=	=	=	=	=
Ptop	kPa	101.3	=	=	=	=	=
Pbottom	kPa	111.6	=	=	=	=	=
Diameter	m	9.8	=	=	=	=	=
Condenser duty	MW	0	=	=	=	=	=
Reboiler duty	MW	0	=	=	=	=	=
Absorber Water Wash							
Water rate	L/s	7.3	=	=	=	=	=
Water temperature	C	37.8	=	=	=	=	=
Temperature	C	48.1	=	=	=	=	=
Pressure	kPa	101.3	=	=	=	=	=

"=" indicates a value equal to the base case, Case 1.

Table 3-1. Summary of Process Simulation Inputs (continued)

Description	Unit	Value					
		Case 1	Case 2	Case 3	Case 4	Case 1b	Case 2b
Rich Amine Pump							
Discharge pressure	kPa	482	=	689	689	=	=
Efficiency	%	65	=	=	=	=	=
Rich/Lean Exchanger							
Rich outlet temperature	C	112.8	=	=	=	=	=
Rich outlet pressure	kPa	345	=	551	551	=	=
Stripper							
Packing type	-	Same as Absorber	=	=	=	=	=
Packing height	MW	9.6	=	=	=	=	=
Ptop	kPa	192.3	=	58.8	58.8	=	=
P, inter1	kPa	N/A	N/A	283.6	283.6	N/A	N/A
Pbottom	kPa	202.6	=	=	=	=	=
Diameter	m	5.5	=	3.7 to 4.9	3.7 to 4.9	5.8	5.8
Reboiler duty	MW	500	499	390	391	572	572
Reflux Condenser							
Temperature	C	35	N/A	N/A	35	=	N/A
Pressure	kPa	202.5	N/A	N/A	405.1	=	N/A
Lean Amine Pump							
Discharge pressure	kPa	447.9	=	=	=	=	=
Efficiency	%	65	=	=	=	=	=
Compression							
Discharge pressure	kPa	13884	=	=	=	=	=
N stages (excludes CO_2 pump)	-	4	9	8	5	=	9
Compressor polytropic efficiency	%	79.5	=	=	=	=	=
CO_2 pump efficiency	%	60	=	=	=	=	=
Cooler outlet temperatures	C	40, 35 (Stages 1-4, Pump)	130 (Stages 1-8); 40,35 (Stage 9, Pump)	130 (Stages 3-6); 40, 35 (Stages 7-8, Pump)	40, 35 (Stages 3-5, Pump)	=	130 (Stages 1-8); 40, 35 (Stage 9, Pump)
Cooler pressure drops	kPa	13.8	=	=	=	=	=

"=" indicates a value equal to the base case, Case 1.

3.2 Process Simulation Results

The process simulation flow diagrams, process simulation results summary, and material balances are given in the following subsections.

3.2.1 Process Simulation Flow Diagrams

The following four figures present simplified process flow diagrams for Cases 1 through 4, respectively. The flow diagrams from Case 1 and 1b are identical, as are the ones for Case 2 and 2b, hence there are only four flow diagrams for the six cases studied. The single compressor train has multiple stages, interstage coolers, and separators that are not all shown on the diagram for clarity. Similarly, there are four parallel amine absorber and regenerator trains that are shown as one train on the diagram.

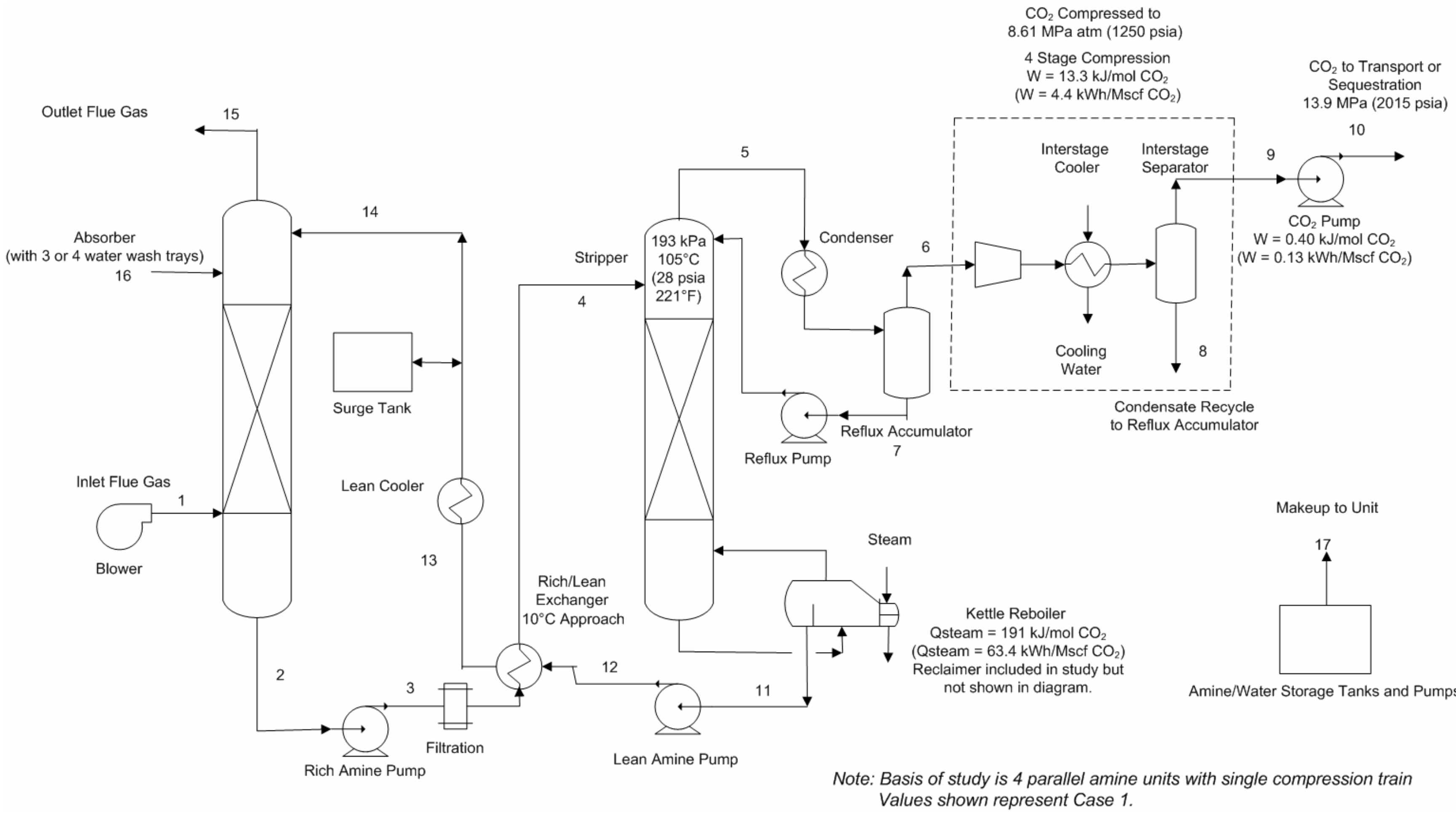

Figure 3-1. Simple Stripper Configuration – No Integration of Compression Heat with MEA Regeneration (Cases 1 and 1b)

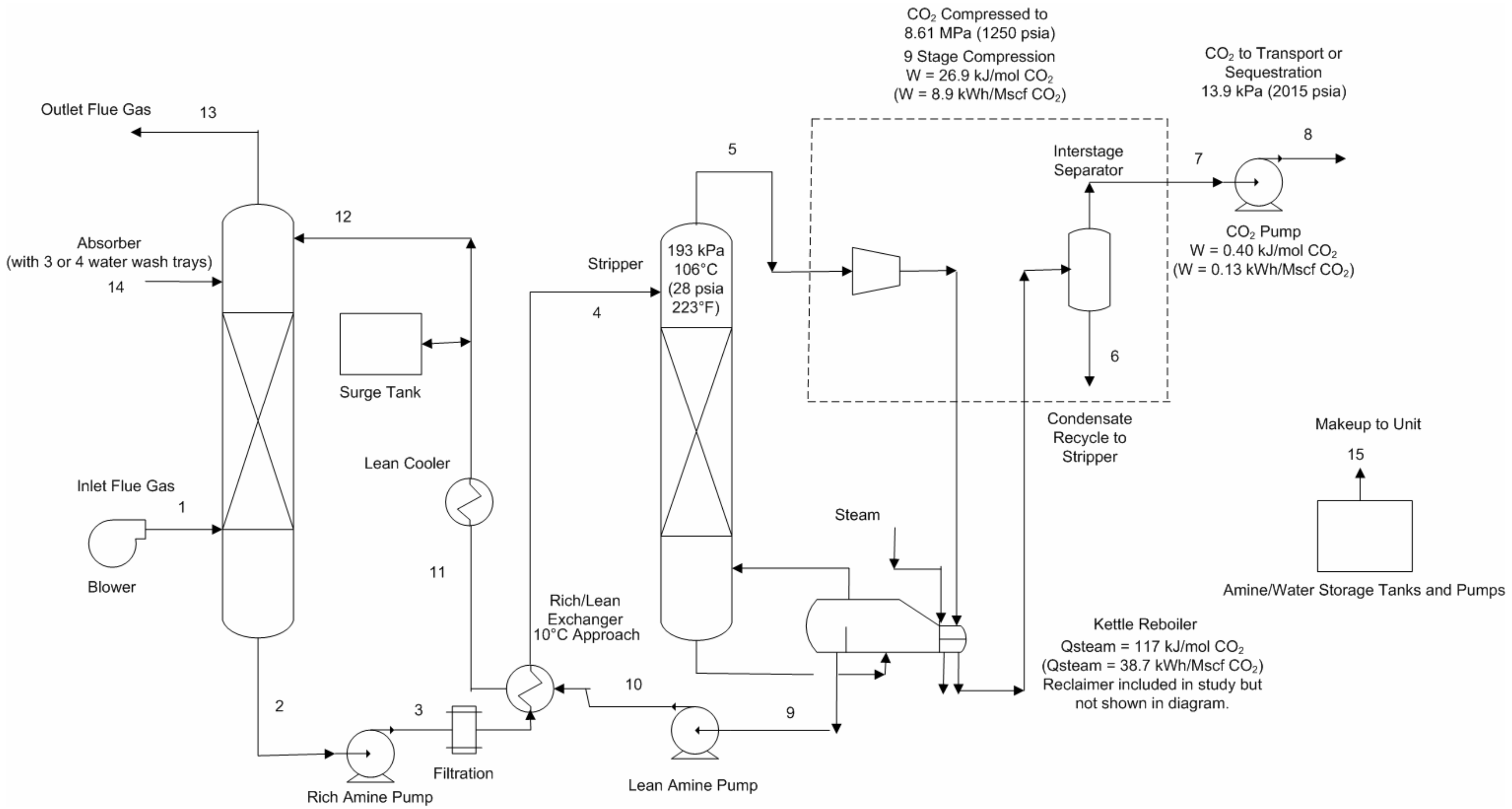

Figure 3-2. Vapor Recompression with Heat Recovery (Cases 2 and 2b)

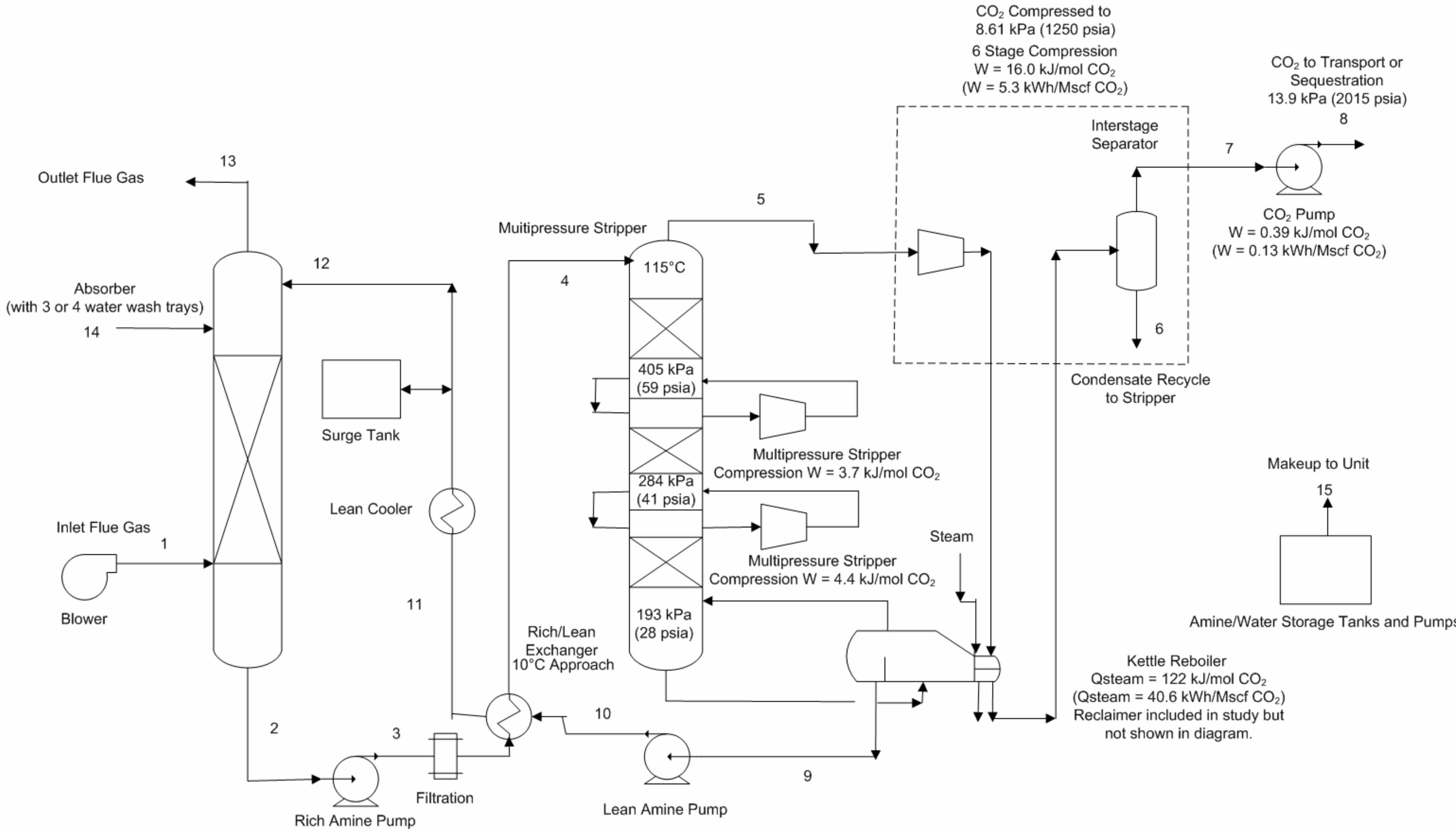

Figure 3-3. Multipressure stripping with Vapor Recompression Heat Recovery (Case 3)

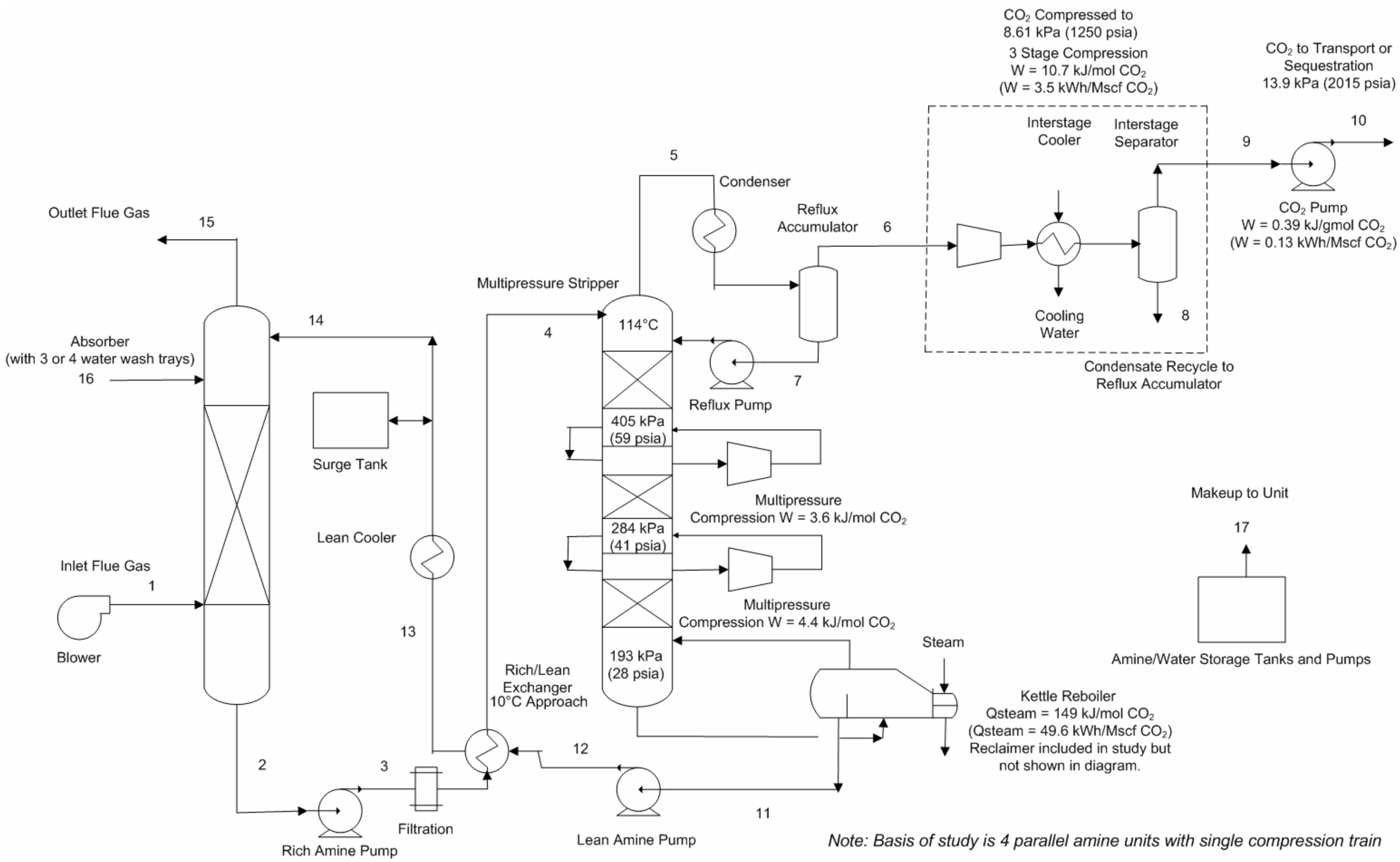

Figure 3-4. Multipressure stripping without Vapor Recompression Heat Recovery (Case 4)

3.2.2 Summary of Process Simulation Results

The process simulation results are summarized in the following table. For each of the cases, the key simulation parameters (e.g. amine circulation rates, reboiler duties, and compression power) are given. All cases use the same lean amine loading. The number of compressor stages excludes the final CO_2 pump that increases the discharge pressure to near 13.9 MPa (2000 psig).

Table 3-2. Summary of Process Simulation Results

		Case 1	Case 2	Case 3	Case 4	Case 1b	Case 2b
CO_2 Percent Removal	%	90	90	90	90	95	95
Amine circulation rate (lean)	L/s	2,798	2,794	2,854	2,857	3,347	3,339
Rich amine CO_2 loading	gmol/L	1.96	1.96	1.94	1.94	1.85	1.85
Lean amine CO_2 loading	gmol/L	1.18	1.18	1.18	1.18	1.18	1.18
Rich/lean heat exchanger duty	MW	510	510	513	513	562	562
Reflux condenser duty	MW	161.7	-	-	73.3	199.0	-
Gross reboiler duty	MW	500	499	390	391	572	572
Net reboiler duty	MW	500	305	320	391	572	336
Lean cooler duty	MW	287	288	301	301	395	396
Rich amine pump power	kW	1,741	1,741	2,332	2,332	2,081	2,081
Lean amine pump power	kW	1,160	1,159	1,163	1,164	1,389	1,387
CO_2 compressor stages required	-	4	9	8	5	4	9
CO_2 compressor power	kW	34,845	70,515	62,411	48,634	36,777	77,847
CO_2 pump power	kW	1,001	993	995	993	1,049	1,048

* Excludes CO_2 pump stage

Comparison of Case 1 with Case 1b and of Case 2 with 2b reveals the requirements for increasing CO_2 removal from 90% to 95% for cases with regeneration reflux (Case 1 and 1b) and without reflux (Cases 2 and 2b). The trends with and without reflux are similar. To improve the CO_2 removal, both cases require an amine circulation rate increase of 20% and a corresponding total amine pump power increase of 20%. Absolute increases in heat exchanger duties are similar with and without reflux.

However, compression power increases about 6% with reflux and increases 10% without reflux.

Comparison of Case 1 with Case 2 and of Case 1b with Case 2b indicates the effect of refluxing some of the regenerator stripper overheads. In cases where reflux is not used, some of the latent heat is recovered in the reboiler, resulting in lower reboiler duties. This advantage is partially offset by the higher compression power requirements that nearly double for the combined total of compressor and CO_2 pump power. The effects of overhead reflux are less pronounced for the multipressure stripping; the total compression power for Case 3, without reflux, is 28% greater than the Case 4 total, rather than double the value. This result occurs because the multipressure stripper incorporates some internal reflux between each of the distinct pressure segments.

Comparison of Case 1 with Case 4 reveals the effect of using multipressure stripping with reflux. Comparison of Case 2 with Case 3 reveals the effect of using multipressure stripping without reflux. Multipressure stripping does not significantly alter the amine circulation rate. Compression power increases by 39% with reflux (Cases 1 and 4), but it decreases by 11% for scenarios without reflux (Cases 2 and 3). Reboiler heat duty decreases by 22% regardless of the reflux option. Rich amine pump power requirements are higher for multipressure stripping because the rich amine enters the high pressure stripper section.

3.2.3 Material Balances

Material balances for each of the six cases are given in the following series of tables. Each material balance gives the stream composition, flow rate, temperature, pressure, vapor fraction, density, and average molecular weight. The stream numbers at the top of the table correspond to flow diagrams presented in Section 3.2.1.

Table 3-3. Material Balance for Case 1.

	1	2	3	4	5	6	7	8	9
Mole Flow kgmol/hr									
H2O	7982.2	353944.9	353942.4	352687.5	12390.5	276.0	12114.4	245.8	30.2
CO2	10459.2	2.9	2.9	326.1	9432.3	9420.9	11.3	1.1	9419.8
MEA	0.0	5753.2	5756.5	7800.8	8.5	0.0	8.5	0.0	0.0
N2	62339.5	1.2	1.2	1.2	1.2	1.2	0.0	0.0	1.2
O2	4046.3	0.1	0.1	0.1	0.1	0.1	0.0	0.0	0.1
MEA+	0.0	20812.3	20811.5	20345.3	0.0	0.0	0.0	0.0	0.0
MEACOO-	0.0	18375.8	18373.3	16795.2	0.0	0.0	0.0	0.0	0.0
HCO3-	0.0	2063.4	2066.6	3464.6	0.0	0.0	0.0	0.0	0.0
CO3--	0.0	184.4	183.7	40.6	0.0	0.0	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.3	0.3	0.3	0.0	0.0	0.0	0.0	0.0
HCOO-	0.0	4.0	4.0	4.0	0.0	0.0	0.0	0.0	0.0
Total Flow kgmol/hr	84827	401142	401142	401466	21833	9698	12134	247	9451
Total Flow kg/hr	2479931	10070083	10070083	10070083	638888	419621	219267	4476	415145
Total Flow m^3/hr	2074927	10568	10567	10889	353063	121361	221	4	721
Temperature C	55.0	57.9	58.0	112.6	105.0	35.0	35.0	--	35.0
Pressure kPa	111.3	101.3	482.4	206.7	192.4	192.4	192.4	--	8611.9
Vapor Frac	1	0	0	0	1	1	0	--	1
Liquid Frac	0	1	1	1	0	0	1	1	0
Density kg/m^3	1.20	952.91	952.95	924.75	1.81	3.46	993.17	1000.32	576.14
Average MW	29.24	25.10	25.10	25.08	28.63	43.27	18.04	18.00	43.9

Table 3-3. Material Balance for Case 1 (continued)

	10	11	12	13	14	15	16	17
Mole Flow kgmol/hr								
H2O	30.2	354976.7	354976.4	355010.2	355329.8	8463.6	511.6	0.0
CO2	9419.8	25.7	25.7	19.3	0.0	1037.0	0.0	0.0
MEA	0.0	23342.8	23343.1	23293.8	22706.8	0.0	0.0	0.1
N2	1.2	0.0	0.0	0.0	0.0	62338.3	0.0	0.0
O2	0.1	0.0	0.0	0.0	0.0	4046.1	0.0	0.0
MEA+	0.0	11210.7	11210.8	11219.9	11467.2	0.0	0.0	0.0
MEACOO-	0.0	10379.1	10378.9	10419.0	10758.7	0.0	0.0	0.0
HCO3-	0.0	774.9	775.2	738.6	191.0	0.0	0.0	0.0
CO3--	0.0	25.7	25.7	28.4	255.8	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	1.3	1.3	1.3	2.0	0.0	0.0	0.0
HCOO-	0.0	4.0	4.0	4.0	4.0	0.0	0.0	0.0
Total Flow kgmol/hr	9451	400741	400741	400735	400715	75885	512	0
Total Flow kg/hr	415145	9647442	9647442	9647442	9647442	2073899	9216	4
Total Flow m^3/hr	664	10075	10052	10019	9502	1996514	9	0
Temperature C	51.1	122.6	122.7	67.8	40.0	48.1	37.8	37.8
Pressure kPa	13883.6	202.5	447.9	337.7	137.8	101.3	172.3	172.3
Vapor Frac	1	0	0	0	0	1	0	0
Liquid Frac	0	1	1	1	1	0	0	1
Density kg/m^3	625.36	957.54	959.78	962.94	1015.30	1.04	993.93	1511.17
Average MW	43.9	24.28	24.28	24.28	24.08	27.33	18.02	61.08

Table 3-4. Material Balance for Case 2.

	1	2	3	4	5	6	7
Mole Flow kgmol/hr							
H2O	7982.2	353944.9	353942.4	352687.5	14285.2	14260.8	24.3
CO2	10459.2	2.9	2.9	326.0	9454.3	30.0	9424.2
MEA	0.0	5753.2	5756.5	7800.7	8.8	8.8	0.0
N2	62339.5	1.2	1.2	1.2	1.2	0.0	1.2
O2	4046.3	0.1	0.1	0.1	0.1	0.0	0.1
MEA+	0.0	20812.3	20811.5	20345.4	0.0	0.0	0.0
MEACOO-	0.0	18375.8	18373.3	16795.2	0.0	0.0	0.0
HCO3-	0.0	2063.4	2066.6	3464.6	0.0	0.0	0.0
CO3--	0.0	184.4	183.7	40.6	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.3	0.3	0.3	0.0	0.0	0.0
HCOO-	0.0	4.0	4.0	4.0	0.0	0.0	0.0
Total Flow kgmol/hr	84827	401142	401142	401466	23750	14300	9450
Total Flow kg/hr	2479931	10070083	10070083	10070083	674006	258773	415233
Total Flow m^3/hr	2074927	10568	10567	10889	385144	259	718
Temperature C	55.0	57.9	58.0	112.6	106.2	--	35.0
Pressure kPa	111.3	101.3	482.4	206.7	192.4	--	8611.9
Vapor Frac	1	0	0	0	1	--	1
Liquid Frac	0	1	1	1	0	1	0
Density kg/m^3	1.20	952.91	952.95	924.75	1.75	1000.32	578.71
Average MW	29.24	25.10	25.10	25.08	28.63	18	43.9

Table 3-4. Material Balance for Case 2 (continued)

	8	9	10	11	12	13	14	15
Mole Flow kgmol/hr								
H2O	24.3	355354.9	355354.6	355388.3	355707.8	8463.6	505.6	0.0
CO2	9424.2	25.6	25.6	19.3	0.0	1037.0	0.0	0.0
MEA	0.0	23348.8	23349.2	23300.0	22713.1	0.0	0.0	0.1
N2	1.2	0.0	0.0	0.0	0.0	62338.3	0.0	0.0
O2	0.1	0.0	0.0	0.0	0.0	4046.1	0.0	0.0
MEA+	0.0	11207.7	11207.6	11216.7	11464.1	0.0	0.0	0.0
MEACOO-	0.0	10376.0	10375.7	10415.8	10755.3	0.0	0.0	0.0
HCO3-	0.0	774.9	775.2	738.7	191.0	0.0	0.0	0.0
CO3--	0.0	25.7	25.7	28.5	255.9	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	1.3	1.3	1.3	2.0	0.0	0.0	0.0
HCOO-	0.0	4.0	4.0	4.0	4.0	0.0	0.0	0.0
Total Flow kgmol/hr	9450	401119	401119	401113	401093	75885	506	0
Total Flow kg/hr	415233	9654099	9654099	9654099	9654099	2073899	9109	4
Total Flow m^3/hr	668	10059	10059	10026	9509	1996514	9	0
Temperature C	50.6	122.6	122.6	67.8	40.0	48.1	37.8	37.8
Pressure kPa	13883.6	202.5	447.9	344.6	137.8	101.3	172.3	172.3
Vapor Frac	1	0	0	0	0	1	0	0
Liquid Frac	0	1	1	1	1	0	0	1
Density kg/m^3	621.99	959.79	959.79	962.95	1015.31	1.04	993.93	1511.17
Average MW	43.9	24.28	24.28	24.28	24.08	27.33	0.00	61.08

Table 3-5. Material Balance for Case 3.

	1	2	3	4	5	6	7	8
Mole Flow kgmol/hr								
H2O	7982.2	361732.2	361728.5	360469.9	5784.7	5759.1	25.5	25.5
CO2	10459.2	2.9	2.9	309.8	9446.2	23.9	9422.3	9422.3
MEA	0.0	6191.2	6196.2	8210.4	3.9	3.9	0.0	0.0
N2	62339.5	1.2	1.2	1.2	1.2	0.0	1.2	1.2
O2	4046.3	0.1	0.1	0.1	0.1	0.0	0.1	0.1
MEA+	0.0	21055.5	21054.3	20605.7	0.0	0.0	0.0	0.0
MEACOO-	0.0	18672.8	18669.0	17103.5	0.0	0.0	0.0	0.0
HCO3-	0.0	2009.8	2014.7	3415.1	0.0	0.0	0.0	0.0
CO3--	0.0	184.3	183.1	41.4	0.0	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.3	0.3	0.3	0.0	0.0	0.0	0.0
HCOO-	0.0	4.1	4.1	4.1	0.0	0.0	0.0	0.0
Total Flow kgmol/hr	84827	409854	409854	410161	15236	5787	9449	9449
Total Flow kg/hr	2479931	10279873	10279873	10279873	520214	105046	414035	414035
Total Flow m^3/hr	2075515	10786	10785	11115	119606	105	716	666
Temperature C	55.0	58.6	58.8	112.6	115.4	--	35.0	50.6
Pressure kPa	111.3	101.3	689.1	413.5	405.1	--	8611.9	13883.6
Vapor Frac	1	0	0	0	1	--	1	1
Liquid Frac	0	1	1	1	0	1	0	0
Density kg/m^3	1.19	953.05	953.12	924.89	4.35	1000.03	578.55	621.82
Average MW	29.24	25.18	25.08	25.06	34.29	18.00	43.91	43.91

Table 3-5. Material Balance for Case 3. (continued)

	9	10	11	12	13	14	15
Mole Flow kgmol/hr							
H2O	363062.5	363062.2	363096.7	363423.2	8462.2	505.4	0.0
CO2	26.1	26.2	19.7	0.0	1037.7	0.0	0.0
MEA	23859.7	23860.2	23809.9	23210.1	0.0	0.0	0.1
N2	0.0	0.0	0.0	0.0	62338.3	0.0	0.0
O2	0.0	0.0	0.0	0.0	4046.1	0.0	0.0
MEA+	11452.8	11452.7	11462.0	11714.9	0.0	0.0	0.0
MEACOO-	10603.1	10602.7	10643.7	10990.6	0.0	0.0	0.0
HCO3-	791.8	792.1	754.8	195.2	0.0	0.0	0.0
CO3--	26.3	26.3	29.1	261.5	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	1.3	1.3	1.4	2.1	0.0	0.0	0.0
HCOO-	4.1	4.1	4.1	4.1	0.0	0.0	0.0
Total Flow kgmol/hr	409828	409828	409821	409802	75884	505	0
Total Flow kg/hr	9864109	9864109	9864109	9864109	2073902	9106	4
Total Flow m^3/hr	10280	10280	10246	9718	1997060	9	0
Temperature C	122.6	122.6	68.3	40.0	48.1	37.8	37.8
Pressure kPa	202.5	447.9	344.6	137.8	101.3	172.3	172.3
Vapor Frac	0	0	0	0	1	0	0
Liquid Frac	1	1	1	1	0	0	1
Density kg/m^3	959.52	959.52	962.68	1015.03	1.04	993.65	1510.74
Average MW	24.15	24.15	24.15	24.08	27.33	0.00	61.08

Table 3-6. Material Balance for Case 4.

	1	2	3	4	5	6	7	8
Mole Flow kgmol/hr								
H2O	7982.2	361732.2	361728.5	360469.9	5210.7	139.1	5071.6	113.6
CO2	10459.2	2.9	2.9	309.8	9432.3	9422.8	9.5	0.9
MEA	0.0	6191.2	6196.2	8210.5	3.9	0.0	3.9	0.0
N2	62339.5	1.2	1.2	1.2	1.2	1.2	0.0	0.0
O2	4046.3	0.1	0.1	0.1	0.1	0.1	0.0	0.0
MEA+	0.0	21055.5	21054.3	20605.7	0.0	0.0	0.0	0.0
MEACOO-	0.0	18672.8	18669.0	17103.4	0.0	0.0	0.0	0.0
HCO3-	0.0	2009.8	2014.7	3415.1	0.0	0.0	0.0	0.0
CO3--	0.0	184.3	183.1	41.4	0.0	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.3	0.3	0.3	0.0	0.0	0.0	0.0
HCOO-	0.0	4.1	4.1	4.1	0.0	0.0	0.0	0.0
Total Flow kgmol/hr	84827	409854	409854	410161	14648	9563	5085	114
Total Flow kg/hr	2479931	10279873	10279873	10279873	509261	417237	92024	2085
Total Flow m^3/hr	2074927	380754	380727	392362	4186084	59281	93	2
Temperature C	55.0	58.6	58.8	112.6	114.0	35.0	35.0	--
Pressure kPa	111.3	101.3	689.1	413.5	405.1	405.1	405.1	--
Vapor Frac	1	0	0	0	1	1	0	--
Liquid Frac	0	1	1	1	0	0	1	1
Density kg/m^3	1.20	953.32	953.39	925.16	4.45	7.04	991.50	994.55
Average MW	29.24	25.08	25.08	25.06	34.29	43.63	18.06	18.06

Table 3-6. Material Balance for Case 4 (continued)

	9	10	11	12	13	14	15	16	17
Mole Flow kgmol/hr									
H2O	25.5	25.5	362931.1	362930.7	362965.2	363291.6	8462.2	505.4	0.0
CO2	9421.9	9421.9	26.2	26.2	19.7	0.0	1037.7	0.0	0.0
MEA	0.0	0.0	23860.5	23860.8	23810.5	23210.9	0.0	0.0	0.1
N2	1.2	1.2	0.0	0.0	0.0	0.0	62338.3	0.0	0.0
O2	0.1	0.1	0.0	0.0	0.0	0.0	4046.1	0.0	0.0
MEA+	0.0	0.0	11452.4	11452.3	11461.6	11714.4	0.0	0.0	0.0
MEACOO-	0.0	0.0	10602.8	10602.5	10643.5	10990.3	0.0	0.0	0.0
HCO3-	0.0	0.0	791.5	791.9	754.6	195.1	0.0	0.0	0.0
CO3--	0.0	0.0	26.3	26.3	29.1	261.4	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.0	1.3	1.3	1.4	2.1	0.0	0.0	0.0
HCOO-	0.0	0.0	4.1	4.1	4.1	4.1	0.0	0.0	0.0
Total Flow kgmol/hr	9449	9449	409696	409696	409690	409670	75884	505	0
Total Flow kg/hr	415152	415152	9861721	9861721	9861721	9861721	2073902	9106	4
Total Flow m^3/hr	715	667	10289	10275	10241	9713	70496213	9	0
Temperature C	35.0	50.6	122.6	122.7	68.3	40.0	48.1	37.8	37.8
Pressure kPa	8611.9	13883.6	202.5	447.9	344.6	137.8	101.3	172.3	172.3
Vapor Frac	1	1	0	0	0	0	1	0	0
Liquid Frac	0	0	1	1	1	1	0	1	1
Density kg/m^3	580.31	622.63	958.49	959.80	962.96	1015.32	1.04	993.93	1511.17
Average MW	43.91	43.91	24.15	24.15	24.15	24.08	27.33	18.02	61.08

Table 3-7. Material Balance for Case 1b

	1	2	3	4	5	6	7	8
Mole Flow kgmol/hr								
H2O	7982.2	423667.6	423665.1	422357.4	15265.4	291.4	14974.0	259.4
CO2	10459.2	3.1	3.1	260.4	9957.3	9943.2	14.0	1.2
MEA	0.0	9000.0	9003.3	10970.3	11.8	0.0	11.8	0.0
N2	62339.5	1.4	1.4	1.4	1.4	1.4	0.0	0.0
O2	4046.3	0.2	0.2	0.2	0.2	0.2	0.0	0.0
MEA+	0.0	23531.9	23531.1	23129.2	0.0	0.0	0.0	0.0
MEACOO-	0.0	21203.7	21201.2	19636.1	0.0	0.0	0.0	0.0
HCO3-	0.0	1937.2	1940.5	3392.9	0.0	0.0	0.0	0.0
CO3--	0.0	192.8	192.1	47.5	0.0	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.5	0.5	0.4	0.0	0.0	0.0	0.0
HCOO-	0.0	4.8	4.8	4.8	0.0	0.0	0.0	0.0
Total Flow kgmol/hr	84827	479543	479543	479800	25236	10236	15000	261
Total Flow kg/hr	2479931	11980513	11980556	11980556	713989	442890	271099	4725
Total Flow m^3/hr	2074927	12541	12541	12912	409999	128091	273	5
Temperature C	55.0	62.2	62.3	112.6	106.9	35.0	35.0	--
Pressure kPa	111.3	101.3	482.4	206.7	192.4	192.4	192.4	--
Vapor Frac	1	0	0	0	1	1	0	--
Liquid Frac	0	1	1	1	0	0	1	1
Density kg/m^3	1.20	955.30	955.34	927.84	1.74	3.46	993.17	1000.32
Average MW	29.24	24.98	24.98	24.97	27.66	43.27	18.04	18.00

Table 3-7. Material Balance for Case 1b (continued)

	9	10	11	12	13	14	15	16	17
Mole Flow kgmol/hr									
H2O	31.9	31.9	424458.5	424458.1	424498.5	424191.0	8369.1	418.8	0.0
CO2	9942.1	9942.1	30.7	30.7	23.1	0.0	519.2	0.0	0.0
MEA	0.0	0.0	27917.9	27918.1	27859.2	27162.7	0.1	0.0	0.1
N2	1.4	1.4	0.0	0.0	0.0	0.0	62338.1	0.0	0.0
O2	0.2	0.2	0.0	0.0	0.0	0.0	4046.1	0.0	0.0
MEA+	0.0	0.0	13399.9	13399.9	13410.8	13709.4	0.0	0.0	0.0
MEACOO-	0.0	0.0	12406.1	12405.8	12453.8	12863.4	0.0	0.0	0.0
HCO3-	0.0	0.0	925.9	926.4	882.7	228.0	0.0	0.0	0.0
CO3--	0.0	0.0	30.8	30.7	34.0	305.4	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.0	1.5	1.5	1.6	2.4	0.0	0.0	0.0
HCOO-	0.0	0.0	4.8	4.8	4.8	4.8	0.0	0.0	0.0
Total Flow kgmol/hr	9975	9975	479176	479176	479168	478467	75273	419	0
Total Flow kg/hr	438166	438166	11535318	11535318	11535318	11523751	2049403	7545	4
Total Flow m^3/hr	759	704	12052	12018	11979	11349	1980445	8	0
Temperature C	35.0	50.6	122.6	122.7	72.2	40.0	48.1	37.8	37.8
Pressure kPa	8611.9	13883.6	202.5	447.9	337.7	137.8	101.3	172.3	172.3
Vapor Frac	1	1	0	0	0	0	1	0	0
Liquid Frac	0	0	1	1	1	1	0	1	1
Density kg/m^3	577.11	621.99	957.10	959.80	962.96	1015.39	1.03	993.93	1511.17
Average MW	43.9	43.9	24.29	24.29	24.29	24.08	27.23	18.02	61.08

Table 3-8. Material Balance for Case 2b.

	1	2	3	4	5	6	7	8
Mole Flow kgmol/hr								
H2O	7982.2	423667.6	423665.1	422357.3	17623.2	17597.5	25.6	25.6
CO2	10459.2	3.1	3.1	260.4	9975.7	32.6	9943.0	9943.0
MEA	0.0	9000.0	9003.3	10970.2	12.1	12.1	0.0	0.0
N2	62339.5	1.4	1.4	1.4	1.4	0.0	1.4	1.4
O2	4046.3	0.2	0.2	0.2	0.2	0.0	0.2	0.2
MEA+	0.0	23531.9	23531.1	23129.2	0.0	0.0	0.0	0.0
MEACOO-	0.0	21203.7	21201.2	19636.1	0.0	0.0	0.0	0.0
HCO3-	0.0	1937.2	1940.5	3392.9	0.0	0.0	0.0	0.0
CO3--	0.0	192.8	192.1	47.5	0.0	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	0.0	0.5	0.5	0.4	0.0	0.0	0.0	0.0
HCOO-	0.0	4.8	4.8	4.8	0.0	0.0	0.0	0.0
Total Flow kgmol/hr	84827	479543	479543	479800	27613	17642	9970	9970
Total Flow kg/hr	2479931	11980556	11980556	11980555	757298	319202	438096	438096
Total Flow m^3/hr	2075515	12545	12544	12916	449859	319	759	705
Temperature C	55.0	62.2	62.3	112.6	107.9	--	35.0	50.6
Pressure kPa	111.3	101.3	482.4	206.7	192.4	--	8611.9	13883.6
Vapor Frac	1	0	0	0	1	--	1	1
Liquid Frac	0	1	1	1	0	1	0	0
Density kg/m^3	1.19	955.03	955.07	927.58	1.68	1000.03	576.94	621.82
Average MW	29.24	24.98	24.98	24.97	27.66	18.00	43.9	43.9

Table 3-8. Material Balance for Case 2b. (continued)

	9	10	11	12	13	14	16
Mole Flow kgmol/hr							
H2O	424656.9	424656.5	424696.9	425078.8	8369.1	412.5	0.0
CO2	30.6	30.7	23.1	0.0	519.2	0.0	0.0
MEA	27917.0	27917.6	27858.7	27157.1	0.1	0.0	0.1
N2	0.0	0.0	0.0	0.0	62338.1	0.0	0.0
O2	0.0	0.0	0.0	0.0	4046.1	0.0	0.0
MEA+	13400.3	13400.2	13411.1	13706.8	0.0	0.0	0.0
MEACOO-	12406.2	12405.7	12453.7	12859.6	0.0	0.0	0.0
HCO3-	926.2	926.6	882.9	228.3	0.0	0.0	0.0
CO3--	30.8	30.7	34.0	305.9	0.0	0.0	0.0
H3O+	0.0	0.0	0.0	0.0	0.0	0.0	0.0
OH-	1.5	1.5	1.6	2.4	0.0	0.0	0.0
HCOO-	4.8	4.8	4.8	4.8	0.0	0.0	0.0
Total Flow kgmol/hr	479374	479374	479367	479344	75273	413	0
Total Flow kg/hr	11538887	11538887	11538887	11538887	2049403	7431	4
Total Flow m^3/hr	12026	12026	11986	11368	1981006	7	0
Temperature C	122.6	122.7	72.2	40.0	48.1	37.8	37.8
Pressure kPa	202.5	447.9	344.6	137.8	101.3	172.3	172.3
Vapor Frac	0	0	0	0	1	0	0
Liquid Frac	1	1	1	1	0	0	1
Density kg/m^3	959.52	959.53	962.68	1015.03	1.03	993.65	1510.74
Average MW	24.29	24.29	24.29	24.08	27.23	0.00	61.08

References (Section 3)

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4.0 EQUIPMENT SIZING AND SELECTION

This section describes the general approach used to size and select the equipment in the CO_2 capture and compression system for this study. As discussed earlier in Section 2, the modeling results were scaled to a 500 MW unit using guidelines from DOE (McGurl, 2004) on the coal composition, plant heat rates, and fuel heating value. The scaled heat and material balances served as the basis for the design of the full-scale plant. A combination of spreadsheet calculations and simulation tools (Aspen Plus, DesignII, and PDQ$) were used to help size the equipment in the process. The basis of the study was four parallel amine units followed by a common downstream compression system.

The general approach in selecting and sizing the equipment in the process was first to use equipment that is considered "standard" to most MEA unit designs and CO_2 compression systems as well as to investigate the possibility of using new approaches in key areas to help reduce overall costs. It is important to note that some of these alternative equipment types may help reduce the overall cost of the process but do not impact the case-by-case comparison results for reducing the parasitic energy demand on the unit since the equipment selections are common to all cases.

The key assumptions used to size the equipment are discussed in the subsections below. A summary table comparing the size requirements and type of equipment for each case is provided at the end of this section.

4.1 Flue Gas Blower

The cool flue gas from the wet FGD scrubber is pressurized using a blower before it enters the absorber. The flue gas enters at the bottom of the absorber and flows upward countercurrent to the amine flow. Thus, it needs to overcome a substantial pressure drop (typically 10.3-17.2 kPa or 1.5-2.5 psi) as it passes through the absorber column with 50 feet of packing material. A 75% efficiency factor was used for each of the blowers in the process trains.

4.2 Absorber

This is the vessel where the MEA-based sorbent contacts the flue gas and absorbs CO_2. The cross sectional area of the absorber is determined from the flue gas flow rate and a design flux of 0.08 kmol/m^2-s (0.02 lbmol/ft^2-s) as provided in the design basis for the study. A maximum practical diameter of 9.7 meters (32 feet) was chosen since this is the upper limit of the costing software used in the study and results in four absorber trains and downstream equipment. [Absorber diameters reported in the literature ranged from 7.9 to 12.8 meters (26 to 42 ft) (Rao, 2004).] Since the purpose of this study was to evaluate the economic tradeoffs between the cases and all the cases used the same absorber design and number of trains, the effect of using a larger absorber diameter and possibly fewer numbers of trains was not completed at this time. Optimization of the absorber size and resulting process trains would be completed during detailed design of a commercial application. It may be that larger vessels, although fewer in number, may be more costly due to size restrictions on the materials and overall construction of the vessels.

The absorber is a vertical, packed column with a water wash section at the top to remove vaporized amine from the overhead stream. The height of the packing is approximately 15 meters (50 ft) and was optimized in previous work by the University of Texas (Freugia, 2002). Although tray absorbers have been operated successfully in the field, packed columns tend to allow for reduced pressure drop, increased gas throughput, improved gas contacting efficiency, and reduced potential for foaming. Carbon steel was selected for the vessel and stainless steel was selected for the packing (GPSA, 1998; Chinn).

4.3 Rich Amine Pump

Rich amine solution from the bottom of the absorber is pumped to an elevated pressure to avoid acid gas breakout in the rich/lean exchanger and to overcome the operating pressure and height requirements in the stripper. Discharge pressures vary among the cases since the operating pressure at the top of the stripper in Cases 3 and 4 (405.3 kPa) is higher than with Cases 1 and 2 (202.6 kPa) (59 psi vs. 27 psi, respectively). Approximately 738 L/s (11,700 gpm) of amine solution is pumped per train. A pump efficiency of 65% was used in the study with 50% sparing of equipment. Stainless steel metal components were selected for the pump.

4.4 Filtration

A filtration step is needed to minimize operating problems caused by solids and other contaminants in the amine solution. There is considerable variation from plant to plant regarding the placement of filters (i.e., before or after the regenerator), the fraction of the stream routed to the filter, and the type of filters used (Skinner, 1995). For this study, it was assumed that a slipstream of the circulating amine (typically 10-20%) is filtered to remove suspended solids then sent to an activated carbon bed filter that adsorbs impurities (degradation products of MEA) and other contaminants from the sorbent stream. This filtration step was also assumed to occur on the dirtier rich amine stream although the difference in size and cost would not vary significantly if installed on the lean stream instead. Carbon steel vessels can be used with this application.

Many different types of mechanical filters are commonly used in amine systems, including leaf-type precoat filters, sock filters, canister or cartridge filters. These filters remove iron sulfide particles, which may enter with the gas or result from corrosion within the system, down to 10-25 micron size. In a well-running system, the filters may need to be replaced on a monthly basis. More frequent replacement may be necessary if the amine is especially dirty or severe foaming is an issue. The mechanical filters remove particulate matter but cannot remove heat stable salts, degradation products, chlorides and other soluble contaminants, or hydrocarbons.

Activated carbon beds can remove hydrocarbons (if present in a utility plant setting) and high-molecular weight degradation products. Activated carbon cannot remove heat stable salts and chlorides. Carbon filters generally need at least 15 minutes of contact time and a maximum superficial velocity of four gpm per square foot (Skinner, 1995). Over a period of time (3-6 months) the carbon bed needs to be replaced and the used bed can be sent back to the suppliers or regenerated on site depending on the plant.

4.5 Rich/Lean Exchanger

The rich amine is preheated from 57.8^{o}C (136^{o}F) to about 113^{o}C (235^{o}F) by heat exchange with the hot lean amine (from the regenerator) in a rich/lean amine exchanger prior to being regenerated. These temperatures were based on a 10^{o}C approach on the hot side of the exchanger. Approximately 65% of the available heat is transferred from the hot lean amine to the cooler rich amine. The heat exchanger is operated at elevated pressure to prevent acid gas breakout and to prevent corrosion of the heat exchanger, control valves, and down-stream piping. The shell and tube heat exchanger is operated with the rich amine on the tube side and at low linear velocity (0.6-0.9 m/sec or 2-3 ft/sec) to prevent or minimize erosion and corrosion. Stainless steel was selected for the rich amine tubes and carbon steel for the shell. A heat transfer coefficient of 511 W/m^2-K (90 Btu/hr-ft^2-F) was used along with a mean temperature difference of about 9^{o}C (16^{o}F) to determine the required surface area of the exchanger. Heat transfer coefficients ranging from 426 to 625 W/m^2-K (75 to 110 Btu/hr-ft^2-F) for this service were found in the literature (GPSA, 1998).

Alternately, it may be possible to use plate and frame heat exchangers for this service. Since the plates are generally designed to form channels giving high turbulent flow, the plate and frame heat exchangers produce higher heat transfer coefficients for liquid flow than most other types. The high heat transfer coefficients are developed through the effective use of pressure drop. For large-scale applications such as the one being considered in this study, plate and frame exchangers offer large surface areas and high heat transfer rates in a small volume and at reduced cost. Gasketed plate and frame exchangers could cost anywhere from 10 to 60% of the corresponding shell-and-tube exchangers. However, pressure drop considerations are critical in the design of this type of exchanger. For the purposes of this study, the more conventional shell-and-tube heat exchanger was selected for costing purposes.

It should also be noted that it would reduce the reboiler steam requirements if a lower (5^{o}C) approach could be specified on the hot side. However, this would more than double the surface area required because of the "pinch point" that develops on the cold end of the exchanger. The economic tradeoffs from going from a 10^{o}C approach to a 5^{o}C approach were evaluated briefly and it was determined that the payback (based on steam savings and increased capital costs for the exchanger) would not warrant using the lower approach at this time. The cost analysis was therefore based on using a 10^{o}C approach.

4.6 Regeneration

Regeneration of the rich amine solution involves a stripper column with reflux and reboiler sections. Each of these areas is discussed below.

4.6.1 Stripper

The main function of the amine regenerator is to remove CO_2 from the rich solution by steam stripping. The absorption reactions are reversed with heat supplied by stripping generated in the reboiler. The rich solution flows down through the regenerator, which is a packed column. Steam rising up through the column strips the CO_2 from the amine solution. The height of the packing is approximately 10.7 meters (35 ft) and the diameter of the column is determined based on a conventional 80% approach of flooding in the column. The packing height was optimized in previous work done by the University of Texas (Freugia, 2002). It is important to note that Cases 3 and 4 involve a multipressure stripper that varies in diameter as the operating pressure increases from about 202.6 kPa (29 psi) at the bottom of the column to 405.3 kPa (59 psi) at the top.

The top area of the column above the rich amine feed point acts as reflux to prevent vaporized or entrained amine from being carried overhead. A substantial surge volume is provided in the base of the regenerator (7.6 m^3 or 267 ft^3). A regenerator bottoms pressure of 202.6 kPa (29.4 psia) and a temperature of 123°C (253°F) is sufficient to strip the acid gas from the solution. The stripper design is based on carbon steel and the packing is of stainless steel construction (GPSA, 1998).

4.6.2 Reboiler

A heat source is used in the tube side of the reboiler to vaporize part of the lean amine solution and generate steam for stripping. In Cases 1 and 4, 446.1 kPa (64.7 psia) saturated steam extracted from the power plant is used for reboiler heat. In Cases 2 and 3, heat will be obtained from a combination of steam from the power plant and hot CO_2/water vapor from the compression stages. A kettle type reboiler is used in this study. Solution flows by gravity from the bottom of the regenerator to the kettle reboiler. A weir maintains the liquid level in the reboiler such that the tube bundle is always submerged. Vapor disengaging space is provided in the exchanger. The vapor is piped back to the regenerator column to provide stripping vapor, while bottom product is drawn from the reboiler. Kettle reboilers are relatively easy to control and no two-phase flow or circulation rate considerations are required. Because of the vapor disengagement requirement, kettles are built with a larger shell.

The reboiler tube bundle is of stainless steel construction while the shell can be carbon steel (GPSA, 1998). A heat transfer coefficient of 625 W/m^2-K (110 Btu/hr-ft^2-F) was used to size the reboiler tubes when steam is used as the heat source. Values in the literature are readily available for this service and ranged from 568 to 909 W/m^2-K (100 to 160 Btu/hr-ft^2-F) (GPSA, 1998). When CO_2/water vapor from compression is used as the reboiler heat source, the overall heat transfer coefficient was estimated from 1) the boiling that occurs on the outside of the tube and 2) the gas-side resistance inside the tubes (the effect of water condensation on the heat transfer coefficient was neglected). The heat transfer coefficient varies with the process gas pressure since it comes from different compression stages (GPSA, 1998). The heat transfer coefficients ranged from approximately 199 W/m^2-K to 511 W/m^2-K (35 Btu/hr-ft^2-F to 90 Btu/hr-ft^2-F) at the higher pressures when CO_2 process gas was used as the heat medium. The log mean temperature differences (LMTD) ranged from 17^oC to 47^oC (30^oF to 85^oF).

It should also be noted that the total reboiler duty for Cases 3 and 4 (391 MW or 1,333 MMBtu/hr) were lower than those required Cases 1 and 2 (499 MW or 1,704 MMBtu/hr). This is because some of the heat for stripping is provided from the two stages of intermediate compression in the multipressure column configuration evaluated in Cases 3 and 4.

4.6.3 Reflux

In Cases 1 and 4, the acid gases and steam leave the top of the regenerator and pass through a reflux condenser and reflux drum where most of the steam is condensed, cooled, and separated from the acid gases. The acid gases then proceed to the compression stage of the process. Depending on the case, the reflux condenser cools the stripper overhead stream from 108-116.1^oC (226-241^oF) to about 35^oC (95^oF) based on availability of cooling water at 29.4^oC (85^oF). The temperature of the return cooling water from the condensers was limited to 43.3^oC (110^oF) to avoid potential scaling problems. A heat transfer coefficient of 454 W/m^2-K (80 Btu/hr-ft^2-F) was used to size the exchanger; other literature sources (GPSA, 1998) consider 397 to 511 W/m^2-K (70 to 90 Btu/hr-ft^2-F) to be typical for this service. Stainless steel was selected for the tube side of the exchanger and carbon steel was selected for the shell. Air-cooled exchangers could be used but are not the preferred choice due to the large heat requirements for this application and resulting size of the coolers.

The reflux drum collects the condensed steam, which is pumped to the top of the regenerator column. The reflux drum was sized using the DesignII process simulator assuming a horizontal vessel with a 5-minute residence time. Stainless steel material was used. Cases 2 and 3 do not utilize a reflux system.

4.6.4 Reclaimer

A reclaimer system was included in the study to remove high boiling degradation products and sludge. In such a system, a small slipstream of the MEA solution in circulation (3%) is taken from the solution leaving the reboiler and fed to a small, steam-heated kettle or reclaimer. The reclaimer operates at the pressure of the stripper column. This allows the reclaimer vapor product to be used directly for reboiling the still. Therefore, there is no energy penalty for the heat requirement of this process. At the start of the MEA reclaimer cycle, the feed to the reclaimer boils near the regenerator bottoms temperature of 116 to 127^{o}C (240-260^{o}F). As the non-volatile impurities collect in the reclaimer, the temperature rises. The reclaimer cycle is generally stopped when the temperature in the reclaimer reaches between 138^{o}C and 140^{o}C (280 to 300^{o}F). The bottom sludge (reclaimer waste) is sent for disposal.

4.7 Lean Amine Pump

Lean amine solution from the bottom of the amine regenerator is pumped to an elevated pressure to overcome the pressure drop in the rich/lean amine exchanger and lean amine cooler and flow to the top of the absorber. The lean loading for all of the cases were optimized and the minimum total work was achieved when the lean loading is 0.25 mol CO_2/mol MEA. The lean amine circulation rate is about 700 L/s (11,100 gpm) and the discharge pressure is at least 446.1 kPa (64.7 psia). A pump efficiency of 65% was used with 50% sparing of equipment.

4.8 Surge Tank

The surge tank for the lean amine solution was sized based on a 30-minute residence time. Carbon steel was selected for the surge tank.

4.9 Lean amine cooler

After the rich/lean amine exchanger, the lean amine must be further cooled in a solution cooler (or trim cooler) before it is pumped back into the absorber column. The solution cooler lowers the lean amine temperature from approximately 67^oC (153^oF) to 40^oC (104^oF) using cooling water in a counter-current, shell and tube exchanger (assuming 29.4^oC or 85^oF cooling water is available on site and is heated to 43.3^oC or 110^oF). Higher temperatures can result in excessive amine evaporative loss and decreased acid gas absorption effectiveness. Since the amine solution passing through the tubes is lean and has had most of the CO_2 removed, carbon steel tubes can be used as well as for the shell of the exchanger. A heat transfer coefficient of 483 W/m^2-K (85 Btu/hr-ft^2-F) was used to size the exchanger. Literature values (GPSA, 1998) indicate the heat transfer coefficient for this service could be on the order of 454 to 511 W/m^2-K (80 to 90 Btu/hr-ft^2-F.)

4.10 Makeup Amine/Water

Because of the vaporization losses it is usually necessary to add make-up amine and water to maintain the desired solution strength. The frequency depends on a number of factors including the heat source in the reboiler and temperature of the reflux condenser. In addition to vaporization, losses of the amine solution may also occur from degradation, entrainment, and mechanical sources. All of the amine entering the stripper does not get regenerated. Flue gas impurities (oxygen, sulfur oxides and nitrogen dioxide) react with the amine to form heat stable salts and reduce the absorption capacity of the amine. Although upstream SO_x and NO_x units were assumed to be used in this study to minimize the amount of contaminants entering the amine unit, the nominal loss of MEA was conservatively estimated at 1.5 kg MEA/tonne CO_2 based on a review of the literature (Rao, 2004). There are only minor differences in the evaporative losses among the cases since the condensate from the vapor recompression cases will be recycled back to the amine unit. If this was not the case, the evaporative MEA losses for the vapor recompression cases with no reflux would be much higher. The amine makeup tank was sized to hold one month's worth of chemical and the makeup water about one day.

4.11 Compressors

The CO_2 compression equipment and the approach for selecting and sizing it are described below.

- *Compression Process Equipment*. The CO_2 from the amine unit is compressed in a single train to 8.6 MPa (1250 psia) and then pumped with multistage centrifugal pumps to 13.9 Mpa (2015 psia) pipeline pressure. The efficiency for this type of pump is 60%.

- *Axial versus Centrifugal Compression for First Stage.* The total CO_2 capture flow rate for the 500 MW base case is approximately 2,025 m^3/min (71,500 acfm). For this size range, either a small axial compressor or a large centrifugal compressor could be used (according to compressor selection guidance in the Gas Processors Suppliers Association manuals). Axial compressors are expected to be similar in cost to centrifugals and may even be somewhat higher since they are not as widely used in industry. The efficiency of an axial compressor is approximately the same as that of a multistage centrifugal compressor (79.5% polytropic efficiency) for this application. Given the lack of any apparent cost or efficiency advantages, and the complexities of maintaining and operating different compressor types with differing maintenance schedules, centrifugal compressors were used in all of the cases.

- *Compression Stages for Various Cases.* The number of compression stages was determined based on a temperature limit and/or compression ratio depending on the case being evaluated in the CO_2 Capture study. The number of compression stages was based on a 177^{o}C (350^{o}F) maximum temperature limit and maximum compression ratio of 3.

4.12 Compressor Drivers

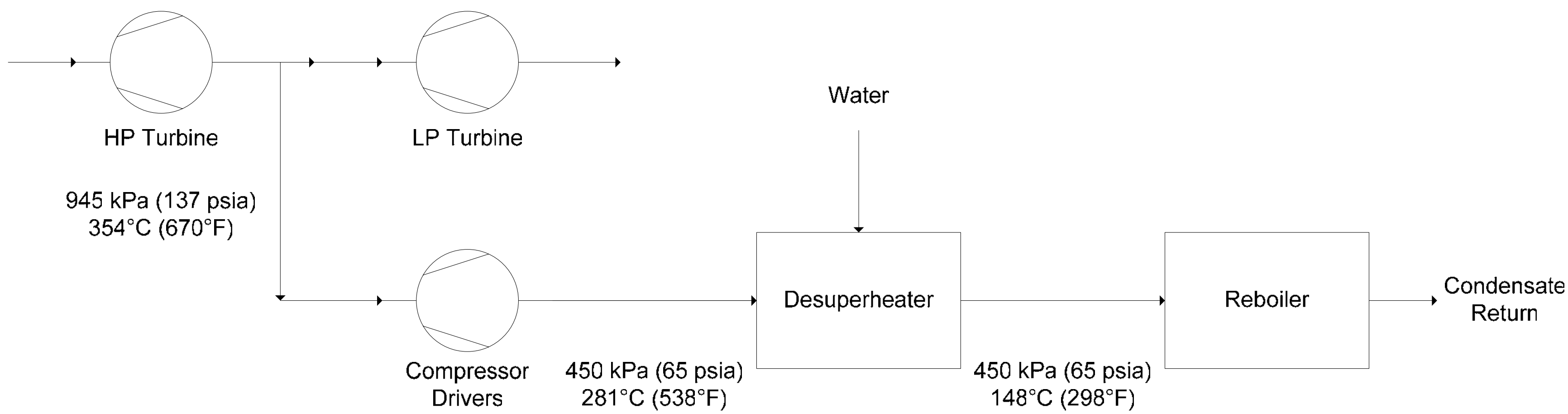
The decision to use steam or electric drivers for the compressors is directly related to the overall strategy for heat integration. If one assumes a constant power output from the power plant, it is necessary to bring in new boiler capacity and power generation dedicated to the operation of the CO_2 capture equipment. An alternate approach is to hold the heat input to the power plant constant, and de-rate the power generation capacity. This approach has been used in other recent published studies and is the approach that has been taken in this analysis.

As shown in Figure 4-1, superheated steam is taken from the power plant at an intermediate pressure of 944.6 kPa and 355^oC (137 psia and 670 ˚F) to provide the necessary reboiler heat for each of the cases. This steam is used to drive the compressor train with a steam turbine, where the steam pressure drops to 446.1 kPa (64.7 psia). The resulting steam temperature at 446.1 kPa (65.7 psia) is approximately 281^oC (538^oF) and, as a result, must be desuperheated with water to provide 446.1 kPa (64.7 psia) steam to feed the reboilers. In cases where the amount of steam required for the reboiler is not enough to drive the compressors, the remaining compressor load is provided with an electric motor using electricity produced by the power plant.

4.13 Interstage Coolers

For Cases 1 and 4, water-cooled exchangers were used for interstage compression cooling. The interstage cooler temperature was based on the availability of cooling water at 29.4^oC (85^oF) and a CO_2 temperature of 40^oC (104^oF) on the tube side of the exchanger. The CO_2 stream is cooled to 35^oC (95^oF) prior to pumping from 8.6 MPa (1250 psia) to 13.9 MPa (2015 psia). Cooling water flow to the intercoolers is done in parallel. The exchanger shell and tubes are made of carbon steel and stainless steel, respectively.

Figure 4-1. Low Pressure Steam Integration from Power Plant



For the vapor recompression cases, the hot CO_2/water vapor stream from the compressor discharge is sent to the tubes of the reboiler to provide additional heat. This stream is cooled to 130°C (266°F) with the amine solution in the reboiler. (The last two stages, however, are cooled with cooling water to 40°C, or 104°F, to facilitate the higher pressure compression and 35°C, or 95°F, for pumping to the final sequestration pressure of 13.9 MPa.) Some of the water vapor condenses from the stream and flows to the downstream separator for removal. The separated gas passes to the next stage of compression. The tube bundles and piping require stainless steel construction due to the corrosive environment with CO_2 and condensing water. Thus, for the vapor recompression cases, the interstage coolers are actually additional tube bundles in the kettle reboiler and not separate exchanger vessels as with Cases 1 and 4 where water cooling is used. The tube bundles are of stainless steel material. The heat transfer coefficients for the interstage coolers were calculated as discussed in Section 4.5.2 for the vapor recompression scenario. The same heat transfer coefficients were used for the water coolers.

4.14 Interstage Separators

Separators are required to remove the condensed liquids from the compression interstage coolers. The separators were sized with DesignII as horizontal vessels with a 5-minute liquid residence time. The sizing calculations are based on general principles that take into account gravity settling for separating the liquid and gas phases and can be used as a preliminary estimate of the size requirements for the separators. The vessels are of stainless steel construction.

4.15 Cooling Water System

A cooling water system is included to provide the necessary cooling for the various cases. A mechanical draft evaporative cooling tower is used with cooling water return and supply temperatures of 43°C to 29°C (110°F to 85°F). The flow rate to the cooling tower varies depending on the case. Cases 1 and 4 require the most cooling water since they utilize cooling water in the lean amine cooler, reflux condenser, and compression interstage coolers (8,800 and 7,400 L/s or 139,000 and 117,000 gpm, respectively). Cases 2 and 3 require less cooling water since they do not have a reflux condenser and use vapor recompression in the interstage coolers (6,100 and 6,400 L/s or 96,000 and 106,000 gpm, respectively). Circulating cooling water pumps are included at approximately 24 meters (80 ft) head.

4.16 Equipment Not Included in Study

When the absorber is operated at higher pressures, as is common in gas-treating applications, the pressure of the rich amine is typically reduced in a flash tank causing a fraction of the absorbed hydrocarbons and acid gases to be removed from solution prior to the amine stripper. For this application, the inlet flue gas is at low pressure and an amine flash tank will not be needed.

A flue gas cooler was not included in this study. Since the gas is coming from an FGD unit, it should already be cooled before entering the amine CO_2 capture equipment.

Dehydration of the CO_2 product stream is dependent on its end-use. If the CO_2 is going to be used in a local enhanced oil recovery field or aquifer for sequestration, dehydration would most likely not be necessary. On the other hand, if the CO_2 were to be transported in a long pipeline, then it would be necessary to dehydrate the CO_2 stream. The cost of dehydration would be the same among the cases studied and is rather small in comparison to the overall costs of CO_2 capture and compression. For these reasons, dehydration of the CO_2 product stream was not included in the cost analysis.

4.17 Equipment Comparison for Cases

Table 4-1 shows a comparison of the equipment size requirements for the various cases in this study. The table shows the major equipment used in each case along with a brief description of the key sizing parameters. The main differences between Cases 1 through 4 are discussed below.

Minor Equipment Differences from Base Case:

- The absorber is the same size for all cases because it is based on gas flow rate;
- Small differences in rich/lean pump flow rates and discharge pressures (Cases 3 and 4 require higher discharge pressures from the rich amine pump since the multipressure stripper operates at 405.3 kPa or 59 psia at the top of the column rather than 202.6 kPa or 27 psia for Cases 1 and 2). The differences in flow rate result in minor variations in filtration requirements;
- The stripper column diameter gets smaller from bottom to top as the pressure changes in the multistripper cases where there is significantly less vapor. Cases 1 and 2 require about 5.5 meters (18 ft) diameter while Cases 3 and 4 varies from 3.7 to 4.9 meters (12 to 16 ft).
- Minor differences in flow rates and temperatures cause slight variation in rich/lean exchanger and lean amine cooler size requirements.
- Minor differences in surge tank capacity due to slight flow rate variations and the makeup and amine tanks are similar in size for all cases.
- Essentially the same CO_2 pump size is required for each case.

More Significant Equipment Differences from Base Case:

- The majority of differences in the equipment size requirements from the base case (Case 1) occur between the interactions of the reflux system and reboiler duty requirements, compression interstage cooling requirements, and compression work for the various flow schemes.
- The reboiler duty for Case 1 is 500 MW or 1705 MMBtu/hr (all steam heat). The reboiler duty for Case 4 is lowered to 391 MW or 1333 MMBtu/hr (all steam heat) because some of the heat requirement is obtained from the multipressure stripper compressors. Case 3 has a lower reboiler steam requirement (320 MW or 1091 MMBtu/hr) due to the heat obtained from the multipressure stripper compressors and the downstream vapor recompression heat recovery. The reboiler steam requirement for Case 2 is lowered to 305 MW (1042 MMBtu/hr) strictly from the downstream vapor recompression heat recovery. Since the same heat transfer coefficient and mean temperature difference is used for the steam reboilers, the differences in reboiler duty

correlate directly to the exchanger size requirements (steam heat contribution only).

- Case 4, which utilizes a multipressure stripper configuration, has a smaller reflux condenser, compression interstage coolers and separators than the base case (Case 1).

- Case 2, which does not employ a reflux system, has a higher compressor interstage condenser duty requirement than the base case. In the base case (Case 1), the stripper overhead stream is first cooled in the reflux condenser and then fed to the compressors such that the bulk duty requirement occurs in the reflux step. The interstage compression coolers for Case 2 require more surface area than the combined reflux and interstage coolers with Case 1 because the heat transfer coefficients and mean temperature differences are lower than those with the reflux cooling. The compressor interstage duty for Case 2 is used to heat the reboiler and is actually represented by additional tube bundles in the reboiler rather than separate exchangers as with Cases 1 and 4.

- Case 3 also does not use a reflux system but is operated at higher pressure in the stripper column. There are two compressors that are used to increase the pressure in the stripper column from 202.6 kPa to 405.3 kPa (27 psia to 59 psia); the discharge from these compressors is not cooled and instead provides some of the heat needed in the stripper to regenerate the solution. Thus, the compressor interstage duty for Case 3 is significantly less than the Case 1 reflux and compressor interstage duty and results in smaller exchanger surface areas even though they are not as efficient heat transfer providers. As with Case 2, the hot compressor process gas is used to heat the reboiler through additional tube bundles in the vessel.

- The compressor work requirements for Cases 2 (70,500 kW or 94,600 hp) and 3 (62,400 kW or 83,700 hp) that utilize vapor recompression and/or multipressure stripper compression are significantly larger than the base case (34,900 kW or 46,700 hp) to meet the temperature and compression ratio requirements. Case 4 also has a higher compressor horsepower requirement (48,600 kW or 65,200 hp), resulting from the two multipressure compressor stages.

- Differences in the separator vessel sizes are a reflection of the different flow configurations as discussed above. In some intercompression stages, condensed liquid will not form; however, separators were included as a safety measure as conditions may vary (i.e., startup/shutdown variations in operation).

Other:

- The cooling water system requirements are largest for Cases 1 and 4 since they use cooling water not only for the lean cooler (common to all cases) but also the reflux condenser and interstage compressor exchangers.

- Cases 1b and 2b were generated to evaluate the possibility of running the CO_2 capture unit at a higher control efficiency (95% instead of 90%) for 95% of the time and then turning the system off for 5% of the time during times of peak electricity demand. As a result, these cases require larger equipment than their counterparts because more CO_2 is removed from the flue gas. Other trends will be the same as discussed above.

Table 4-1. Equipment Sizing Comparison

No.	Description	Units	Case I	Case 2	Case 3	Case 4	Case Ib	Case 2b	Notes on Differences:
	Summary of Case	--	Base Case (no integration of compression with MEA regeneration)	Vapor recompression with heat recovery	Multipressure stripping with vapor recompression heat recovery	Multipressure stripping without vapor recompression heat recovery	Base Case (no integration of compression with MEA regeneration)	Vapor recompression with heat recovery	
	CO_2 capture	%	90	90	90	90	95	95	
1	Flue gas blower (1 of 4 trains) - Flow	kg/hr	620,000	620,000	620,000	620,000	620,000	620,000	
	Pressure drop	kPa	10.3	10.3	10.3	10.3	10.3	10.3	Typical range of 10.3 to 17.2 kPa
2	Absorber (1 of 4 trains) - Diameter	m	10	10	10	10	10	10	Same absorber size required in all cases.
	Packing height	m	15	15	15	15	15	15	
3	Rich amine pump (1 of 4 trains) - Flow	L/s	732	732	744	744	877	877	Small differences in flow rate and discharge pressures. Case 3 and 4 operate at higher stripper pressures.
	Discharge head	m	74	74	92	92	74	74	
	Work	kW	433	433	582	582	522	522	
4	Filtration (1 of 4 trains) - Flow rate	L/s	110	110	112	112	132	132	Very minor differences based on flow through filter.
5	Stripper (1 of 4 trains) - Diameter	m	6	6	4 to 5	4 to 5	6	6	Differences in vessel diameter and operating pressure. In Case 3 and case 4, the vessel diameter changes through the column because it operates at multiple pressures.
	Packing height	m	11	11	11	11	11	11	
6	Reflux condenser (1 of 4 trains) - Flow to condenser	actual m^3/min	1,470	na	na	476	1,708	na	No reflux in Cases 2 and 3. Higher flow in Case 1 than with multipressure Case 4.
	Duty	MW	40	na	na	18	50	na	
	Heat transfer coefficient	W/m^2-K	454	na	na	454	454	na	
	LMTD	C	23	na	na	26	24	na	Cooling water used from 29.4C to 43.3C
	Area	m^2	3,818	na	na	1,579	4,599	na	
7	Reflux accumulator (1 of 4 trains) - Diameter	m	1	na	na	1	2	na	No reflux in Cases 2 and 3.
	Length	m	5	na	na	5	7	na	
8	Reflux pump (1 of 4 trains) - Flow rate	L/s	15	na	na	6	1	na	No reflux in Cases 2 and 3. More condensate in Case 1 than with multipressure Case 4.
	Discharge pressure	kPa	345	na	na	345	345	na	
	Work	kW	7	na	na	4	11	na	
9	Reboiler (1 of 4 trains) - Duty	MW	125	76	80	98	143	84	The reboiler in Cases 1 and 4 are steam heated only.
	Heat transfer coefficient	W/m^2-K	625	625	625	625	625	625	The reboiler in Cases 2 and 3 are heated with steam and vapor recompression. Only the steam portion is included here. See interstage cooler requirements below for vapor recompression portion.
	LMTD	C	26	26	26	26	26	26	
	Area	m^2	7,609	4,654	4,942	6,039	8,742	5,128	
10	Reclaimer (1 of 4 trains) - Duty	MW	31	31	32	32	37	37	Slipstream of 3% used in sizing.
	Heat transfer coefficient	W/m^2-K	625	625	625	625	625	625	
	LMTD	C	12	12	12	12	12	12	
	Area	m^2	4,136	4,139	4,229	4,228	4,945	4,947	
11	Rich/lean amine heat exchanger (1 of 4 trains) - Duty	MW	128	128	128	128	141	141	Slightly differences in duty and LMTDs between cases based on where recycled water is introduced into system.
	Heat transfer coefficient	W/m^2-K	511	511	511	511	511	511	
	LMTD	C	10	10	10	10	10	9	
	Area	m^2	25,284	25,279	25,697	25,920	27,611	27,611	

Table 4-1. Equipment Sizing Comparison (continued)

12	Surge tank (1 of 4 trains) - Volume	L	1,187,733	1,256,999	1,284,629	1,213,850	1,418,240	1,502,645	Slight differences in flow rates between cases; sized on fixed residence time.
13	Lean amine pump (1 of 4 trains) -- Flow	L/s	700	700	713	713	839	833	Lean pumps have about same head but slightly different flow rates.
	Head	m	74	74	74	74	74	74	
	Work	kW	291	291	291	291	350	343	
14	Lean amine cooler (1 of 4) - Duty	MW	72	72	75	75	99	99	Differences in duties and minor changes in LMTDs.
	Heat transfer coefficient	W/m^2-K	483	483	483	483	483	483	
	LMTD	C	17	17	17	17	18	18	
	Area	m^2	8,993	9,012	9,300	9,300	11,251	11,251	
15	Amine storage tank (common to 4 trains) - Volume	L	276,305	276,305	276,305	276,305	291,445	291,445	Based on fixed MEA loss rate (1.5 kg/tonne CO_2)
16	Amine makeup pump (common to 4 trains) - Flow rate	L/s	0.11	0.11	0.11	0.11	0.12	0.12	Continuous basis; likely to be metered in at higher rate over shorter period of time.
17	Water storage tank (common to 4 trains) - Volume	L	628,310	628,310	628,310	628,310	628,310	628,310	
18	Water pump (common to 4 trains) - Flow rate	L/s	7	7	7	7	7	7	
19	CO2 compressors (common to 4 trains) - Flow to pipeline compression	actual m^3/min	2,022	6,417	1,991	988	2,135	7,491	Compression to 8.6 MPa.
	Flow in multipressure stripper compression	actual m^3/min	na	na	7,477 & 4,220	7,477 & 4,220	na	na	Two compressors from 202.6 kPa to 283.7 kPa and 283.7 kPa to 405.3 kPa internal to stripper for Cases 3and 4.
	Total Work	kW	34,824	70,543	62,415	48,620	36,763	77,851	
	Stages	--	4	9	8	5	4	9	
	Driver type		steam/elec	elec/steam	elec/steam	steam/elec	steam/elec	elec/steam	
20	Interstage kettle coolers (common to 4 trains) - Vapor Recompression Duty	MW	na	194	71	na	na	236	Shell and tube water coolers for Cases 1 and 4. Additional kettle tube bundles for vapor recompression interchange in reboiler for Cases 2 and 3. CO2 process gas side cooled to 130C with amine solution in reboiler. Heat transfer coefficient and LMTD vary with interstage compression pressure and temperature.
	Interstage shell & tube water coolers (common to 4 trains) - Duty	MW	60	63	68	52	64	67	CO_2 process gas side cooled to 40C with cooling water from 29.4C to 43.3C. Heat transfer coefficient and LMTD vary with interstage compression pressure and temperature.
	Number of coolers	--	4	9	6	3	4	9	The two compression stages for the stripper do not require cooling.
21	Interstage separators (common to 4 trains) - Total volume	Mliters	96	432	128	29	71	507	
	Number of separators	--	4	9	6	3	4	9	The two compression stages for the stripper do not require cooling nor separators.
22	CO2 pump (common to 4 trains) - Flow rate	$MMsm^3d$	5.4	5.4	5.4	5.4	5.7	5.6	
	Work	kW	999	992	992	992	1051	1051	
23	Cooling water system (common to 4 trains) - Flow rate	L/s	8,743	6,043	6,346	7,330	11,323	7,955	Includes cooling tower, fans, basin and pump pit, and circulation pumps
	Work	kW	5444	2983	3430	4325	6562	4027	For fans and motors

References (Section 4)

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5.0 CAPITAL AND OPERATING COSTS

This section describes the approach used to estimate the capital and operating costs for the CO_2 capture and compression process approaches evaluated in this study. The cost methodology is discussed first, followed by a presentation of the results.

5.1 Capital Costs

The purchased equipment costs for the amine unit and downstream compression train were obtained from a combination of vendor quotes and costing software using the size parameters discussed in Section 4. PDQ$ (Preliminary Design and Quoting Service) is a software package that can be used to estimate current purchased equipment costs for chemical process equipment. (The costs are in September 2004 dollars.) The software estimates costs for fabricated equipment and catalog items that are based on vendor information. The list below shows the source of the purchased equipment costs by type.

- Absorber and Stripper – PDQ$
- Pumps (rich/lean, reflux, makeup water and amine) – PDQ$
- Filtration – Vendor quote for similar application
- Pressure vessels (reflux accumulator and interstage compression separators) – PDQ$
- Exchangers (reflux condenser, rich/lean exchanger, lean amine cooler, reboiler and compressor interstage coolers) – PDQ$
- Storage tanks (amine and water) – PDQ$
- CO_2 compressors and drivers – PDQ$ and vendor estimates for select cases
- CO_2 pump – Vendor quote
- Cooling tower system – PDQ$

Tables 5-1 through 5-6 show the major equipment list and purchased equipment costs for the various cases. The major differences in cost are related to the cost of the compressors and steam/electric drivers and the tradeoffs in where the heat exchange in the process takes place. In general, the reflux condensers required less surface area for heat transfer than the compression interstage coolers because of their higher heat transfer coefficients. The same is true with the reboiler, where straight steam requires less heat transfer area than when using the process gas from the vapor recompression interstages.

The installed costs for purchased equipment (everything but compression) was estimated using typical factors for percentage of purchased cost as reported in chemical engineering literature (Peters and Timmerhaus, 1980). The installed cost factor for compression was based on vendor recommendations for this type of application. Table 5-7 shows the total process plant cost (PPC) for the different cases.

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 6% to account for engineering and home office expenses. A project contingency of 30% was used since the level of project definition seemed to fall in the AACE Estimate Class 3 for budget authorization (McGurl, 2004). A process contingency of 5% was used for all of the cases since the technology is a commercial process and this same factor was used by EPRI in other CO_2 capture studies (McGurl, 2004; EPRI, 2000). An interest and adjustment factor of 10% of the PPC was used to arrive at the total plant investment (TPI); this factor was also similar to other EPRI work in the CO_2 capture area (EPRI, 2000).

Per the Quality Guidelines for Energy System Studies document (McGurl, 2004), the total capital requirement (TCR) is the total of the total plant investment and:

- Prepaid royalties – 0.5% of PPC for new technology and capital charge;
- Startup costs – 2% of TPI and 30 days of variable O&M (discussed in Section 5.2);
- Spare parts – 0.5% of TPC
- Working capital – 30 days of fixed O&M (discussed in Section 5.2); and
- Land – 1% of TPI per literature (Peters and Timmerhaus, 1980).

Table 5-1. Major Equipment List for Base Case 1, 90% Removal
Simple Stripper Configuration -- No Integration of Compression Heat with MEA Regeneration

Equipment No.	Description	Type	Design Condition per Train	Trains	Material of Construction	Purchased Equipment Cost per Train, $	Total Purchased Equipment Cost , $
1	Flue gas blower	Forced draft	620000 kg/hr/ 10.3 to 17.2 kPa (1366000 lb/hr/1.5 to 2.5 psi)	4		510,000	2,040,000
2	Absorber	Packed bed 5-cm (2") rings, 15-m (50-ft) height; water wash section at top (3 or 4 trays); 9.7m (32 ft) diameter	310.3 kPa/149 C (45 psia/300F)	4	CS vessel; SS packing	4,080,000	16,320,000
3	Rich amine pump	Centrifugal	732 L/s @ 76 m (11600 gpm @ 250 ft)	4	SS	68,000	272,000
4	Filtration	Horizontal	791 kPa/ 93C (115 psia/200F)	4	CS	290,000	1,160,000
5	Stripper	Packed bed 5-cm (2") rings, 10-m (33-ft) height; 5.5 m (18 ft) diameter	446 kPa/ 149C (65 psia/300F)	4	CS vessel; SS packing	940,000	3,760,000
6	Reflux condenser	Shell and tube Cooling water	446 kPa/121 C (65 psia/250F) 41 MW (138 MMBtu/hr)	4	SS tubes CS shell	470,000	1,880,000
7	Reflux accumulator	Horizontal	446 kPa/121 C (65 psia/250F)	4	SS	10,000	40,000
8	Reflux pump	Centrifugal	15 L/s @ 446 kPa (242 gpm @ 65 psia)	4	SS	6,000	24,000
9	Reboiler	Horizontal-kettle, 446 kPa (65 psia) steam	619 kPa/ 177C (90 psia/350F) 125 MW (426 MMBtu/hr)	4	Tubes SS; Vessel CS	1,150,000	4,600,000
10	Reclaimer	Horizontal-kettle	30 MW (105 MMBtu/hr)	4	Tubes SS; Vessel CS	620,000	2,490,000
11	Rich/lean amine heat exchanger	Horizontal shell	791 kPa/138C (115 psia/280 F) 127 MW (435 MMBtu/hr)	4	Rich tubes SS; Vessel CS	2,800,000	11,200,000
12	Surge tank	Horizontal vessel		4	CS	109,000	436,000
13	Lean amine pump	Centrifugal	700 L/s @ 76 m (11100 gpm @ 250 ft)	4	SS	68,000	272,000
14	Lean amine cooler	Shell and tube Cooling water	791 kPa/65C (100 psig/150F) 72 MW (245 MMBtu/hr)	4	CS	1,000,000	4,012,000
15	Amine storage tank	Fixed roof tank	291 Mlitre (73 Mgal)	1	CS	58,000	58,000
16	Amine makeup pump	Centifugal		1	CS	1,300	1,300
17	Water storage tank	Fixed roof tank	628 Mlitre (166 Mgal)	1	CS	97,000	97,000
18	Water makeup pump	Centifugal		1	CS	1,900	1,900
19	CO_2 compressors	Multi-stage, centrifugal (202.6 kPa/8.6 MPa) Drivers	34800 kW (46700 hp)	1	SS	7,215,000 8,833,000	7,215,000 8,833,000
20	CO_2 compressor interstage coolers	Shell and tube; water-cooled	59 MW (203 MMBtu/hr)	1	Tubes SS; Vessel CS	749,000	749,000
21	CO_2 compressor separator	Horizontal vessels	96 Mlitre (25 Mgal)	1	SS	621,000	621,000
22	CO_2 pump		8.6 MPa/13.9 Mpa (1250 psia/2015 psia)	1		492,000	492,000
23	Cooling tower system	Includes cooling tower, fans, basin and pump it, and circulation pumps		1		7,797,000	7,797,000
Total Purchased Equipment Costs						37,986,200	74,371,200
Subtotal CO_2 Capture Purchased Equipment Costs (includes cooling tower system)						20,076,200	56,461,200
Subtotal CO_2 Compression Purchased Equipment Costs						17,910,000	17,910,000

Table 5-2. Major Equipment List for Case 2, 90% Removal
Vapor Recompression with Heat Recovery

Equipment No.	Description	Type	Design Condition per Train	Trains	Material of Construction	Purchased Equipment Cost per Train, $	Total Purchased Equipment Cost , $
1	Flue gas blower	Forced draft	620000 kg/hr/ 10.3 to 17.2 kPa (1366000 lb/hr/1.5 to 2.5 psi)	4		510,000	2,040,000
2	Absorber	Packed bed 5-cm (2") rings, 15-m (50-ft) height; water wash section at top (3 or 4 trays); 9.7m (32 ft) diameter	310.3 kPa/149 C (45 psia/300F)	4	CS vessel; SS packing	4,080,000	16,320,000
3	Rich amine pump	Centrifugal	732 L/s @ 76 m (11600 gpm @ 250 ft)	4	SS	68,000	272,000
4	Filtration	Horizontal	791 kPa/ 93C (115 psia/200F)	4	CS	290,000	1,160,000
5	Stripper	Packed bed 5-cm (2") rings, 10-m (33-ft) height; 5.5-m (18 ft) diameter	446 kPa/ 149C (65 psia/300F)	4	CS vessel; SS packing	940,000	3,760,000
6	Reboiler	Horizontal-shell, 446 kPa (65 psia) steam	619 kPa/ 177C (90 psia/350F) 76 MW (261 MMBtu/hr)	4	Tubes SS; Vessel CS	750,000	3,000,000
7	Reclaimer	Horizontal shell	31 MW (105 MMBtu/hr)	4	Tubes SS; Vessel CS	620,000	2,500,000
8	Rich/lean amine heat exchanger	Horizontal shell	791 kPa/138C (115 psia/280F) 127 MMBtu/hr (435 MMBtu/hr)	4	Rich tubes SS; Vessel CS	2,800,000	11,200,000
9	Surge tank	Horizontal vessel		4	CS	109,000	436,000
10	Lean amine pump	Centrifugal	700 L/s @ 76 m (11100 gpm @ 250 ft)	4	SS	68,000	272,000
11	Lean amine cooler	Shell and tube Cooling water	791 kPa/65C (115 psia/150F) 73 MW (246 MMBtu/hr)	4	CS	1,010,000	4,024,000
12	Amine storage tank	Fixed roof tank	291 Mlitre (73 Mgal)	1	CS	58,000	58,000
13	Amine makeup pump	Centifugal		1	CS	1,300	1,300
14	Water storage tank	Fixed roof tank	628 Mlitre (166 Mgal)	1	CS	97,000	97,000
15	Water makeup pump	Centifugal		1	CS	1,900	1,900
16	CO_2 compressors	Multi-stage, centrifugal; 202.6 kPa (29 psia)/8.6 Mpa (1250 psia) Drivers	70500 kW (94600 hp)	1	SS	19,380,000 6,607,000	19,380,000 6,607,000
17	CO_2 compressor interstage coolers	Kettle interchange with reboiler Shell and tube; water cooled	194 MW (662 MMBtu/hr) 63 MW (216 MMBtu/hr)	1	Tubes SS; Vessel CS	5,280,000 669,000	5,280,000 669,000
18	CO_2 compressor separator	Horizontal vessels	432 Mlitre (114 Mgal)	1	SS	778,000	778,000
19	CO_2 pump		8.6 MPa/13.9 Mpa (1250 psia/2015 psia)	1		492,000	492,000
20	Cooling tower system	Includes cooling tower, fans, basin and pump, and circulation pumps		1		4,664,000	4,664,000
Total Purchased Equipment Costs						49,273,200	83,012,200
Subtotal CO_2 Capture Purchased Equipment Costs (includes cooling tower system)						16,067,200	49,806,200
Subtotal CO_2 Compression Purchased Equipment Costs						33,206,000	33,206,000

Table 5-3. Major Equipment List for Case 3, 90% Removal
Multipressure Stripper with Vapor Recompression Heat Recovery

Equipment No.	Description	Type	Design Condition per Train	Trains	Material of Construction	Purchased Equipment Cost per Train, $	Total Purchased Equipment Cost , $
1	Flue gas blower	Forced draft	620000 kg/hr/ 10.3 to 17.2 kPa (1366000 lb/hr/1.5 to 2.5 psi)	4		510,000	2,040,000
2	Absorber	Packed bed 5-cm (2") rings, 15-m (50-ft) height; water wash section at top (3 or 4 trays); 9.7m (32 ft) diameter	310.3 kPa/149 C (45 psia/300F)	4	CS vessel; SS packing	4,080,000	16,320,000
3	Rich amine pump	Centrifugal	744 L/s @ 91 m (11800 gpm @ 300 ft)	4	SS	78,000	312,000
4	Filtration	Horizontal	791 kPa/ 93C (115 psia/200F)	4	CS	300,000	1,200,000
5	Stripper	Packed bed 5-cm (2") rings, 10-m (33-ft) height; multiple diameters (3.6 m to 4.9 m or 12 ft to 16 ft)	446 kPa/ 149C (65 psia/300F)	4	CS vessel; SS packing	630,000	2,520,000
6	Reboiler	Horizontal-shell, 446 kPa (65 psia) steam	619 kPa/ 177C (90 psia/350F) 80 MW (273 MMBtu/hr)	4	Tubes SS; Vessel CS	790,000	3,160,000
7	Reclaimer	Horizontal shell	31 MW (108 MMBtu/hr)	4	Tubes SS; Vessel CS	640,000	2,550,000
8	Rich/lean amine heat exchanger	Horizontal shell	791 kPa/138C (115 psia/280F) 128 MW (438 MMBtu/hr)	4	Rich tubes SS; Vessel CS	2,850,000	11,400,000
9	Surge tank	Horizontal vessel		4	CS	109,000	436,000
10	Lean amine pump	Centrifugal	713 L/s @ 76 m (11300 gpm @ 250 ft)	4	SS	78,000	312,000
11	Lean amine cooler	Shell and tube Cooling water	791 kPa/65C (115 psia/150F) 76 MW (257 MMBtu/hr)	4	CS	980,000	3,926,000
12	Amine storage tank	Fixed roof tank	276 Mlitre (73 Mgal)	1	CS	58,000	58,000
13	Amine makeup pump	Centifugal		1	CS	1,300	1,300
14	Water storage tank	Fixed roof tank	628 Mlitre (166 Mgal)	1	CS	97,000	97,000
15	Water makeup pump	Centifugal		1	CS	1,900	1,900
16	CO_2 compressors	Multi-stage, centrifugal; 203 kPa (29 psia)/284 kPa (41 psia); 284 kPa (41 psia)/405 kPa (59 psia), 405 kPa (59 psia)/8.6 Mpa (1250 psia) Drivers	62400 kW (83700 hp)	1	SS	17,469,000 6,635,000	17,469,000 6,635,000
17	CO_2 compressor interstage coolers	Kettle interchage with reboiler Shell and tube; water cooled	70 MW (241 MMBtu/hr) 68 MW (232 MMBtu/hr)	1	Tubes SS; Vessel CS	2,305,000 687,000	2,305,000 687,000
18	CO_2 compressor separators	Horizontal vessels	128 Mlitre (34 Mgal)	1	SS	420,000	420,000
19	CO_2 pump		8.6 MPa/13.9 Mpa (1250 psia/2015 psia)	1		492,000	492,000
20	Cooling tower system	Includes cooling tower, fans, basin and pump, and circulation pumps		1		5,257,192	5,257,192
Total Purchased Equipment Costs						44,468,392	77,599,392
Subtotal CO_2 Capture Purchased Equipment Costs (includes cooling tower system)						16,460,392	49,591,392
Subtotal CO_2 Compression Purchased Equipment Costs						28,008,000	28,008,000

Table 5-4. Major Equipment List for Case 4, 90% Removal
Multipressure Stripping without Vapor Recompression Heat Recovery

Equipment No.	Description	Type	Design Condition per Train	Trains	Material of Construction	Purchased Equipment Cost per Train, $	Total Purchased Equipment Cost , $
1	Flue gas blower	Forced draft	620000 kg/hr/ 10.3 to 17.2 kPa (1366000 lb/hr/1.5 to 2.5 psi)	4		510,000	2,040,000
2	Absorber	Packed bed 5-cm (2") rings, 15-m (50-ft) height; water wash section at top (3 or 4 trays); 9.7m (32 ft) diameter	310.3 kPa/149 C (45 psia/300F)	4	CS vessel; SS packing	4,080,000	16,320,000
3	Rich amine pump	Centrifugal	744 L/s @ 91 m (11800 gpm @ 300 ft)	4	SS	78,000	312,000
4	Filtration	Horizontal	791 kPa/ 93C (115 psia/200F)	4	CS	300,000	1,200,000
5	Stripper	Packed bed 5-cm (2") rings, 10-m (33-ft) height; multiple diameters (3.6 m to 4.9 m or 12 ft to 16 ft)	446 kPa/ 149C (65 psia/300F)	4	CS vessel; SS packing	630,000	2,520,000
6	Reflux condenser	Shell and tube Cooling water	446 kPa/121C (65 psia/250F) 19 MW (63 MMBtu/hr)	4	SS tubes CS shell	290,000	1,160,000
7	Reflux accumulator	Horizontal	446 kPa/121C (65 psia/250F)	4	SS	10,000	40,000
8	Reflux pump	Centrifugal	6.5 L/s @ 446 kPa (102 gpm @ 65 psia)	4	SS	6,000	24,000
9	Reboiler	Horizontal-shell, 446 kPa (64.7 psia) steam	619 kPa/ 177C (90 psia/350F) 98 MW (333 MMBtu/hr)	4	Tubes SS; Vessel CS	940,000	3,760,000
10	Reclaimer	Horizontal shell	31 MW (108 MMBtu/hr)	4	Tubes SS; Vessel CS	640,000	2,550,000
11	Rich/lean amine heat exchanger	Horizontal shell	791 kPa/138C (115 psia/280F) 128 MW (438 MMBtu/hr)	4	Rich tubes SS; Vessel CS	2,850,000	11,400,000
12	Surge tank	Horizontal vessel		4	CS	109,000	436,000
13	Lean amine pump	Centrifugal	713 L/s @ 76 m (11300 gpm @ 250 ft)	4	SS	70,000	280,000
14	Lean amine cooler	Shell and tube Cooling water	791 kPa/65C (115 psia/150F) 75 MW (256 MMBtu/hr)	4	CS	900,000	3,580,000
15	Amine storage tank	Fixed roof tank	276 Mlitre (73 Mgal)	1	CS	58,000	58,000
16	Amine makeup pump	Centifugal		1	CS	1,300	1,300
17	Water storage tank	Fixed roof tank	628 Mlitre (166 Mgal)	1	CS	97,000	97,000
18	Water makeup pump	Centifugal		1	CS	1,900	1,900
19	CO_2 compressors	psia)/284 kPa (41 psia); 284 kPa (41 psia)/405 kPa (59 psia), 405 kPa (59 psia)/8.6 Mpa (1250 psia) Drivers	48600 kW (65200 hp)	1	SS	13,854,000 7,433,000	13,854,000 7,433,000
20	CO_2 compressor interstage coolers	Shell and tube; water-cooled	52 MW (178 MMBtu/hr)	1	Tubes SS; Vessel CS	330,000	330,000
21	CO_2 compressor separator	Horizontal vessels	29 Mlitre (8 Mgal)	1	SS	205,000	205,000
22	CO_2 pump		8.6 MPa/13.9 Mpa (1250 psia/2015 psia)	1		492,000	492,000
23	Cooling tower system	Includes cooling tower, fans, basin and pump, and circulation pumps		1		6,674,000	6,674,000
Total Purchased Equipment Costs						40,559,200	74,768,200
Subtotal CO_2 Capture Purchased Equipment Costs (includes cooling tower system)						18,245,200	52,454,200
Subtotal CO_2 Compression Purchased Equipment Costs						22,314,000	22,314,000

Table 5-5. Major Equipment List for Case 1b, 95% Removal
Simple Stripper Configuration -- No Integration of Compression Heat with MEA Regeneration

Equipment No.	Description	Type	Design Condition per Train	Trains	Material of Construction	Purchased Equipment Cost per Train, $	Total Purchased Equipment Cost , $
1	Flue gas blower	Forced draft	620000 kg/hr/ 10.3 to 17.2 kPa (1366000 lb/hr/1.5-2.5 psi)	4		510,000	2,040,000
2	Absorber	Packed bed 5-cm (2") rings, 15-m (50-ft) height; water wash section at top (3 or 4 trays); 9.7m (32 ft) diameter	310.3 kPa/149 C (45 psia/300F)	4	CS vessel; SS packing	4,080,000	16,320,000
3	Rich amine pump	Centrifugal	877 L/s @ 76 m (13000 gpm @ 250 ft)	4	SS	76,000	304,000
4	Filtration	Horizontal	791 kPa/ 93C (115 psia/200F)	4	CS	320,000	1,280,000
5	Stripper	Packed bed 5-cm (2") rings, 10-m (33-ft) height; 5.8-m (19-ft) diameter	446 kPa/149C (65 psia/300F)	4	CS vessel; SS packing	1,030,000	4,120,000
6	Reflux condenser	Shell and tube Cooling water	446 kPa/121C (65 psia/250F) 50 MW (170 MMBtu/hr)	4	SS tubes CS shell	540,000	2,160,000
7	Reflux accumulator	Horizontal	446 kPa/121C (65 psia/250F)	4	SS	17,000	68,000
8	Reflux pump	Centrifugal	19 L/s @ 446 kPa (300 gpm @ 65 psig)	4	SS	7,000	28,000
9	Reboiler	Horizontal-shell, 446 kPa (64.7 psia) steam	619 kPa/ 177C (90 psia/350F) 143 MW (488 MMBtu/hr)	4	Tubes SS; Vessel CS	1,300,000	5,200,000
10	Reclaimer	Horizontal-kettle	37 MW (126 MMBtu/hr)	4	Tubes SS; Vessel CS	750,000	2,980,000
11	Rich/lean amine heat exchanger	Horizontal shell	791 kPa/138C (115 psia/280F) 140 MW (479 MMBtu/hr)	4	Rich tubes SS; Vessel CS	3,040,000	12,160,000
12	Surge tank	Horizontal vessel		4	CS	161,000	644,000
13	Lean amine pump	Centrifugal	839 L/s @ 76 m (13300 gpm @ 250 ft)	4	SS	76,000	304,000
14	Lean amine cooler	Shell and tube Cooling water	791 kPa/65C (115 psia/150F) 99 MW (337 MMBtu/hr)	4	CS	1,070,000	4,280,000
15	Amine storage tank	Fixed roof tank	291 Mlitre (77 Mgal)	1	CS	58,000	58,000
16	Amine makeup pump	Centifugal		1	CS	1,300	1,300
17	Water storage tank	Fixed roof tank	628 Mlitre (166 Mgal)	1	CS	97,000	97,000
18	Water makeup pump	Centifugal		1	CS	1,900	1,900
19	CO_2 compressors	Multi-stage, centrifugal; 203 kPa (29 psia)/8.6 Mpa (1250 psia) Drivers	36800 kW (49300 hp)	1	SS	7,341,000 10,039,000	7,341,000 10,039,000
20	CO_2 compressor interstage coolers	Shell and tube; water-cooled	63 MW (217 MMBtu/hr)	1	Tubes SS; Vessel CS	834,000	834,000
21	CO_2 compressor interstage separators	Horizontal vessels	71 Mlitre (19 Mgal)	1	SS	425,000	425,000
22	CO_2 pump		8.6 MPa/13.9 Mpa (1250 psia/2015 psia)	1		508,000	508,000
23	Cooling tower system	Includes cooling tower, fans, basin and pump, and circulation pumps		1		10,020,000	10,020,000
Total Purchased Equipment Costs						42,302,200	81,213,200
Subtotal CO_2 Capture Purchased Equipment Costs (includes cooling tower system)						23,155,200	62,066,200
Subtotal CO_2 Compression Purchased Equipment Costs						19,147,000	19,147,000

Table 5-6. Major Equipment List for Case 2b, 95% Removal
Vapor Recompression with Heat Recovery

Equipment No.	Description	Type	Design Condition per Train	Trains	Material of Construction	Purchased Equipment Cost per Train, $	Total Purchased Equipment Cost , $
1	Flue gas blower	Forced draft	620000 kg/hr/ 10.3 to 17.2 kPa (1366000 lb/hr/1.5 to 2.5 psi)	4		510,000	2,040,000
2	Absorber	Packed bed 5-cm (2") rings, 15-m (50-ft) height; water wash section at top (3 or 4 trays); 9.7m (32 ft) diameter	310.3 kPa/149 C (45 psia/300F)	4	CS vessel; SS packing	4,080,000	16,320,000
3	Rich amine pump	Centrifugal	877 L/s @ 76 m (13900 gpm @ 250 ft)	4	SS	76,000	304,000
4	Filtration	Horizontal	791 kPa/ 93C (115 psia/200F)	4	CS	320,000	1,280,000
5	Stripper	Packed bed 5-cm (2") rings, 10-m (33-ft) height; 5.8-m (19 ft) diameter	446 kPa/149C (65 psia/300F)	4	CS vessel; SS packing	1,030,000	4,120,000
6	Reboiler	Horizontal-shell, 446 kPa (64.7 psia) steam	619 kPa/ 177C (90 psia/350F) 84 MW (286 MMBtu/hr)	4	Tubes SS; Vessel CS	820,000	3,280,000
7	Reclaimer	Horizontal shell	37 MW (126 MMBtu/hr)	4	Tubes SS; Vessel CS	750,000	2,980,000
8	Rich/lean amine heat exchanger	Horizontal shell	791 kPa/138C (115 psig/280F) 140 MW (479 MMBtu/hr)	4	Rich tubes SS; Vessel CS	3,040,000	12,160,000
9	Surge tank	Horizontal vessel		4	CS	161,000	644,000
10	Lean amine pump	Centrifugal	833 L/s @ 76 m (13200 gpm @ 250 ft)	4	SS	76,000	304,000
11	Lean amine cooler	Shell and tube Cooling water	791 kPa/65C (115 psia/150F) 99 MW (337 MMBtu/hr)	4	CS	1,120,000	4,480,000
12	Amine storage tank	Fixed roof tank	291 Mlitre (77 Mgal)	1	CS	58,000	58,000
13	Amine makeup pump	Centifugal		1	CS	1,300	1,300
14	Water storage tank	Fixed roof tank	629 Mlitre (166 Mgal)	1	CS	97,000	97,000
15	Water makeup pump	Centifugal		1	CS	1,900	1,900
16	CO_2 compressors	Multi-stage, centrifugal; 203 kPa (29 psia)/8.6 Mpa (1250 psia) Drivers	77800 kW (104400 hp)	1	SS	20,537,000 7,270,000	20,537,000 7,270,000
17	CO_2 compressor interstage coolers	Kettle interchange with reboiler Shell and tube; water cooled	236 MW (805 MMBtu/hr) 67 MW (228 MMBtu/hr)	1	Tubes SS; Vessel CS	7,127,000 693,000	7,127,000 693,000
18	CO_2 compressor separators	Horizontal vessels	507 Mlitre (134 Mgal)	1	SS	902,000	902,000
19	CO_2 pump		8.6 MPa/13.9 Mpa (1250 psia/2015 psia)	1		508,000	508,000
20	Cooling tower system	Includes cooling tower, fans, basin and pump, and circulation pumps		1		7,180,000	7,180,000
Total Purchased Equipment Costs						56,358,200	92,287,200
Subtotal CO_2 Capture Purchased Equipment Costs (includes cooling tower system)						19,321,200	55,250,200
Subtotal CO_2 Compression Purchased Equipment Costs						37,037,000	37,037,000

Table 5-7. Process Plant Cost for CO2 Capture and Compression Flow Schemes

Parameter	Factor	Units	Case 1	Case 2	Case 3	Case 4	Case 1b	Case 2b
Total CO_2 capture purchased equipment costs (PEC)		$	56,756,000	50,066,000	49,866,000	52,770,000	62,422,000	55,579,000
Purchased equipment installation	18 % of capture PEC	$	10,216,000	9,012,000	8,976,000	9,499,000	11,236,000	10,004,000
Instrumentation and controls (installed)	8 % of capture PEC	$	4,540,000	4,005,000	3,989,000	4,222,000	4,994,000	4,446,000
Piping (installed)	20 % of capture PEC	$	11,351,000	10,013,000	9,973,000	10,554,000	12,484,000	11,116,000
Electrical (installed)	10 % of capture PEC	$	5,676,000	5,007,000	4,987,000	5,277,000	6,242,000	5,558,000
Buildings (including services)	18 % of capture PEC	$	10,216,000	9,012,000	8,976,000	9,499,000	11,236,000	10,004,000
Yard improvements	6 % of capture PEC	$	3,405,000	3,004,000	3,004,000	3,166,000	3,745,000	3,335,000
Service facilities (installed)	24 % of capture PEC	$	13,621,000	12,016,000	11,968,000	12,665,000	14,981,000	13,339,000
Total installed capital for CO_2 capture		$	115,781,000	102,135,000	101,727,000	107,652,000	127,340,000	113,381,000
Total CO_2 compression purchased equipment costs (PEC)			17,911,000	33,207,000	28,008,000	22,314,000	19,147,000	37,037,000
CO_2 compression installed costs	180 % of compression PEC	$	32,240,000	59,773,000	50,414,000	40,165,000	34,465,000	66,667,000
Total process plant cost (PPC)		$	148,021,000	161,908,000	152,141,000	147,817,000	161,805,000	180,048,000
% of Compression train		%	21.8	36.9	33.1	27.2	21.3	37.0
% of CO2 capture (amine unit + cooling system)		%	78.2	63.1	66.9	72.8	78.7	63.0

Table 5-8 shows how the total capital requirement was derived from the process plant cost as described above.

5.2 Operating Costs

The major operating and maintenance (O&M) costs for the CO_2 capture and compression process consist of both fixed cost and variable cost components as shown in Table 5-9. The operating costs are based on a generic site location and should represent a reasonable average of those in various regions of the country.

Table 5-9. Operating and Maintenance Cost Parameters and Values

Fixed O&M Cost Components	**Value**
Total maintenance cost	2.2% of total plant cost
Maintenance cost allocated to labor	12% of maintenance costs
Administration and support labor cost	30% of operating labor
Operating labor	Assume 1 loaded full-time operator
Variable O&M Cost Components	**Value**
MEA cost	$1200/tonne
Water cost	$0.92/1000 gallons
Solid waste disposal cost	$175/tonne waste

Table 5-8. Total Capital Requirement for CO2 Capture Process Flow Schemes

	Factor	Units	Case 1	Case 2	Case 3	Case 4	Case 1b	Case 2b
Process plant cost (PPC) for CO_2 removal and compression			148,021,000	161,908,000	152,141,000	147,817,000	161,805,000	180,048,000
Engineering and home office	6 % of PPC	$	8,881,000	9,714,000	9,128,000	8,869,000	9,708,000	10,803,000
Project contingency	30 % of PPC	$	44,406,000	48,572,000	45,642,000	44,345,000	48,542,000	54,014,000
Process contingency	5 % of PPC	$	7,401,000	8,095,000	7,607,000	7,391,000	8,090,000	9,002,000
Total plant cost (TPC)		$	208,709,000	228,289,000	214,518,000	208,422,000	228,145,000	253,867,000
Interest and inflation adjustment factor	10 % of PPC	$	14,802,100	16,190,800	15,214,100	14,781,700	16,181,000	18,005,000
Total plant investment (TPI)		$	223,511,100	244,479,800	229,732,100	223,203,700	244,326,000	271,872,000
Royalty fees	0.5 % of PPC	$	740,000	810,000	761,000	739,000	809,000	900,000
Startup cost								
-- component 1	2 % of TPI	$	4,470,000	4,890,000	4,595,000	4,464,000	4,887,000	5,437,000
-- component 2	30 days of variable O&M	$	637,000	581,000	587,000	610,000	711,000	641,000
Spare parts	0.5 % of TPC	$	1,044,000	1,141,000	1,073,000	1,042,000	1,141,000	1,269,000
Working capital	30 days of fixed O&M	$	437,000	477,000	449,000	437,000	477,000	530,000
Land	1 % of TPI	$	2,235,000	2,445,000	2,297,000	2,232,000	2,443,000	2,719,000
Total capital requirement (TCR)		$	233,074,100	254,823,800	239,494,100	232,727,700	254,794,000	283,368,000

The fixed O&M cost factors were obtained from the Quality Guidelines for Energy Systems document (except for operating labor). The variable O&M costs were specific to the operation of the CO_2 capture and compression system and depend on the capacity factor (or load factor) of the plant. A capacity factor of 85% was used in the study for two reasons. First, it is the maximum value allowed in the Quality Guidelines for Energy Systems document (McGurl, 2004). Second, the plant is considered to be a "central" baseload plant which will run at full capacity as much as possible. A capacity factor in excess of 90% is not uncommon for these types of plants, and 85% should be a very achievable (if not conservatively low) number. An 85% capacity factor is equivalent to about 7,451 hours per year of operation at full capacity.

The variable O&M components include costs of chemicals consumed, utilities, and services used. The MEA losses were estimated assuming a factor of 1.5 kg MEA/tonne CO_2 based on a review of the literature (Rao, 2004). This includes vaporization losses and degradation of the MEA for electric utility type operations. The vaporization losses do not differ significantly among the cases since the condensed liquids from the stripper overheads are sent back to the amine unit. The cost of MEA reagent is also based on review of available literature (Rao, 2004). The solid waste disposal cost includes such items as activated carbon replacement. An activated carbon bed in the amine circulation path removes some of the compounds formed from the degenerated MEA. These carbon beds need to be replaced, usually every 3-6 months at an estimated consumption rate of about 0.075 kg C/tonne CO_2 and the cost for solid waste disposal is $175/tonne waste (Rao, 2004). A cooling water system is included in the capital costs and so only makeup water requirements are considered as an operating expense. The estimated cost of makeup water is $0.92/1000 gallons (Rao, 2004). The total annual cost for each item is calculated by multiplying the unit cost by the total annual quantity used or consumed and the hours per year, given the plant capacity factor. The fixed and variable O&M costs are shown in Table 5-10.

Table 5-10. Summary of Operating and Maintenance Costs

Parameter	Factor	Units	Case 1	Case 2	Case 3	Case 4	Case 1b	Case 2b
Capacity factor for plant		%	85	85	85	85	85	85
Fixed O&M Costs Total Maintenance Cost	2.2 % of TPC	$/yr	4,592,000	5,022,000	4,719,000	4,585,000	5,019,000	5,585,000
Maintenance cost allocated to labor	12 % of total maintenance cost	$/yr	551,000	603,000	566,000	550,000	602,000	670,000
Administration and support labor cost	30 % of total labor cost	$/yr	24,000	24,000	24,000	24,000	24,000	24,000
Operating labor	1 dedicated operator	$/yr	80,000	80,000	80,000	80,000	80,000	80,000
Total Fixed O&M Costs		$/yr	5,247,000	5,729,000	5,389,000	5,239,000	5,725,000	6,359,000
Variable O&M Costs Reagent cost	1200 $/tonne MEA	$/yr	5,556,000	5,559,000	5,558,000	5,558,000	5,864,000	5,865,000
Water cost	0.92 $/1000 gallon	$/yr	2,052,000	1,372,000	1,441,000	1,719,000	2,620,000	1,787,000
Solid waste disposal cost	175 $/tonne waste	$/yr	41,000	41,000	41,000	41,000	43,000	43,000
Total Variable O&M Costs		$/yr	7,649,000	6,972,000	7,040,000	7,318,000	8,527,000	7,695,000
Total O&M (TOM) Costs		$/yr	12,896,000	12,701,000	12,429,000	12,557,000	14,252,000	14,054,000

The amine CO_2 capture unit and downstream compression also require electricity and steam to operate. However, these utilities are taken into account with the derating of the power plant and, therefore, no explicit cost is associated with them. This approach is the same as that used when describing the operating costs for a utility power plant without CO_2 capture. That is, the cost of electricity includes the fuel costs for the utility boiler but does not explicitly include a cost for the high-pressure steam that is produced in the boiler.

The electrical requirements are a result of the various pumps, fans, etc. in the amine process and cooling tower system plus that needed to supplement the compressor train if the steam drawn from the power plant for the reboiler cannot supply enough work to drive the compression train. The steam taken from the power plant for the reboiler also results in a loss of electrical output from the facility. The low pressure steam at 944.6 kPa (137 psia) and 354.4^{o}C (670^{o}F) is discharged from the low-pressure turbine at about 7.6 kPa (1.1 psia); thus, every 0.45 kg/hr (1 lb/hr) of steam removed from the turbine results in an estimated electric energy penalty of approximately 0.089 kW (0.119 hp). Table 5-11 shows the energy demand and resulting derated power plant capacity for each of the cases. The energy requirements for the base plant, FGD system, ESP, and SCR are also included in the derating shown in the table (about 9% of the gross capacity).

5.3 Annualized Cost Summary

Once the total capital requirement (TCR) and the total O&M costs are known, the total annualized cost of the power plant was estimated as follows.

Total annual revenue requirement, TRR ($/yr) = (TCR * CRF) + TOM

where, TCR =total capital requirement of the power plant, $ and

CRF = capital recovery factor (fraction).

A capital recovery factor of 15% is used in the analysis for the cases. The normalized total capital requirement ($/kW) was also estimated based on the net power generation (derated) at the plant. Table 5-12 shows how these parameters vary for the different cases.

Table 5-11. Derating of CO2 Capture Process Flow Schemes

Parameter	Units	Case 1 Base Case 90%	Case 2 Heat Recovery 90%	Case 3 Multipressure Stripping Heat Recovery, 90%	Case 4 Multipressure Stripping No Heat Recovery, 90%	Case 1b Base Case 95%	Case 2b Heat Recovery 95%
Summary of Total Work and Duty Requirements							
Total compression work	kW (hp)	35000 (47000)	71000 (95000)	62000 (84000)	49000 (65000)	37000 (49000)	78000 (104000)
Total CO2 pump work	kW (hp)	1000 (1300)	1000 (1300)	1000 (1300)	1000 (1300)	1000 (1400)	1000 (1400)
Total reboiler duty	MW (MMBtu/hr)	492.3 (1705)	492.3 (1704)	386.8 (1332)	386.8 (1334)	563.2 (1953)	563.2 (1951)
Reboiler duty from compressor interstage coolers	MW (MMBtu/hr)	0 (0)	187.3 (662)	69.2 (241)	0 (0)	0 (0)	228 (805)
Remaining steam required for reboiler	MW (MMBtu/hr)	492.3 (1705)	305 (1042)	317.7 (1091)	386.8 (1334)	563.2 (1953)	335.2 (1146)
Total rich/lean pump work	kW (hp)	2900 (3900)	2900 (3900)	3500 (4700)	3500 (4700)	3500 (4700)	3500 (4600)
Reflux pump work	kW (hp)	30 (40)	0 (0)	0 (0)	10 (20)	40 (60)	0 (0)
Cooling system	kW (hp)	5400 (7300)	3000 (4000)	3400 (4600)	4100 (5400)	6600 (8800)	4100 (5400)
Flue gas blower	kW (hp)	9200(12000)	9200(12000)	9200(12000)	9200(12000)	9200(12000)	9200(12000)
Derating Results							
Gross capacity	MW	500	500	500	500	500	500
Reboiler heat requirement (from steam)	MW (MMBtu/hr)	492.3 (1705)	305 (1042)	317.7 (1091)	386.8 (1334)	563.2 (1953)	335.2 (1146)
Reboiler steam rate (1)	kg/hr	847,927	518,337	542,440	663,221	971,064	569,674
Steam from turbine for reboiler heat (2)	kg/hr	763,554	466,760	488,464	597,227	874,438	512,989
Steam required for compression (3)	kg/hr	887,530	1,796,055	1,589,643	1,238,742	936,738	1,982,804
Makeup electricity to supplement compression	kW	4867	52189	43233	25186	2446	57706
Corresponding steam loss in power generation (4)	MW	149	91	95	117	171	100
Derated capacity (reboiler only)	MW	351	409	405	383	329	400
Derated capacity (reboiler, compressors, pumps)	MW	327	340	344	340	306	324
Percent reduction in capacity from CO2 capture	%	35	32	31	32	39	35
Total energy penalty (percent of gross) (5)	%	44	41	40	41	48	44
Derated plant capacity CO2 capture and base plant	MW	281	294	298	293	260	278

Notes:
1) Assumes saturated 446.1 kPa (64.7 psia) steam at 147.6C (297.7F) with 913 Btu/lb latent heat available.
2) Prior to water addition to desuperheat from 446.1 kPa (64.7 psia) and 281.2C (538.2F) to saturated conditions.
3) Assumes steam taken from system at 944.6 kPa (137 psia) and 354.4C (670F) to 446.1 kPa (64.7 psia) and 281.2C (538.2F) (with 72% isentropic efficiency, 0.0239 hp-hr/lb or 0.039 kW-hr/kg).
4) Steam turbine operating at 354.4C (670F) and 944.6 kPa (137 psia) to discharge pressure of 7.6 kPa (1.1 psia) produces 0.11896 hp-hr/lb or 0.195 kW-hr/kg at 80% isentropic efficiency (~PRPA conditions).
5) Percent reduction in capacity from base plant is 9% (29 MW for PC, 14 MW for FGD system, 1 MW for ESP, and 3 MW for SCR).

Table 5-12. Total Annual Revenue Requirement and Normalized Capital

Parameter	Units	Case 1	Case 2	Case 3	Case 4	Case 1b	Case 2b
Normalized Total Capital Requirement	$/kW	830	867	805	793	981	1021
Total Annual Revenue Requirement	$/yr	47,857,000	50,925,000	48,353,000	47,466,000	52,471,000	56,559,000

References (Section 5)

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6.0 ECONOMIC ANALYSIS AND RESULTS

This section utilizes the annualized cost summary from Section 5 to develop comparisons between the various cases for cost of electricity and the cost of CO_2 avoidance as well as an evaluation of the strategy of selective operation. Each of these three areas is discussed in more detail below.

6.1 Cost of Electricity

Table 6-1 presents the cost of electricity once CO_2 recovery is added for each of the four 90% recovery cases. The basis for these costs was previously presented in Section 5.

As shown in the table, the cost of electricity is highest ($63.24/MW-hr) for Case 1, which is the conventional MEA system with compression of the CO_2 after the reflux condenser. Case 2, which removes the reflux condenser and incorporates heat recovery, has an electricity cost of $61.81/MW-hr, which represents a 2.2% savings. Case 3, incorporating both heat recovery and multipressure stripping, has the lowest cost at $59.88/MW-hr, resulting in a savings of about 5.2% over Case 1. Finally, Case 4, including multipressure stripping without heat recovery, has a cost of $60.32/MW-hr; this is approximately a 4.6% savings over Case 1. In evaluating the savings in the cost of electricity, the base coal-fired plant costs comprise a significant portion of the overall cost of electricity; this relatively fixed portion makes the cost savings appear smaller than if they were evaluated on just the cost of CO_2 capture (as will be shown in Section 6.2).

Table 6-2 shows the sensitivity of the overall cost of electricity to the assumption about coal-fired power plant costs and the percentage increase in the cost of electricity under each scenario. Although the $25/MW-hr used in Table 6-1 is considered realistic, some utilities will have higher operating costs depending on their location, fuel cost, fuel quality, and other factors. At higher coal-fired plant operating costs, the four cases still remain in the same order, with Case 3 having the lowest cost of electricity. However, the

Table 6-1. Cost of Electricity with CO_2 Removal Equipment Installed					
	No CO2	**Case 1**	**Case 2**	**Case 3**	**Case 4**
Power Plant size, MW	500	500	500	500	500
Net Power Production (after power plant aux. and CO2 capture)	453	281	294	298	293
CO2 Recovery, tonne/hr		415	415	415	415
Power Plant Cost, $/MWh	25				
Annual Power Plant Cost, $/yr	$84,325,950	$84,325,950	$84,325,950	$84,325,950	$84,325,950
CO2 Removal Plant Variable O&M Costs, $/yr		$7,649,000	$6,972,000	$7,040,000	$7,318,000
CO2 Removal Plant Fixed O&M Costs, $/yr		$5,247,000	$5,729,000	$5,389,000	$5,239,000
CO2 Removal Plant Capital Recovery Costs, $/yr		$34,961,000	$38,224,000	$35,924,000	$34,909,000
Total CO2 Removal Costs, $/yr		$47,857,000	$50,925,000	$48,353,000	$47,466,000
Total Power Plant + CO2 Removal Costs, $/yr		$132,182,950	$135,250,950	$132,678,950	$131,791,950
Electricity Costs, $/MW-hr	$/MWh	63.24	61.81	59.88	60.32

Table 6-2. Sensitivity of CO_2-Controlled Cost of Electricity to Coal Plant Cost

	Case 1		**Case 2**		**Case 3**		**Case 4**	
Coal Plant Cost	**$/MW-hr**	**% Inc. COE**	**$/MW-hr**	**% Inc. COE**	**$/MW-hr**	**% Inc. COE**	**$/MW-hr**	**% Inc. COE**
$25/MW-hr	63.24	152.8%	61.81	147.2%	59.88	139.6%	60.32	141.2%
$35/MW-hr	79.38	126.9%	72.22	120.6%	75.10	114.6%	75.76	116.6%
$45/MW-hr	95.52	112.2%	92.63	105.8%	90.32	100.7%	91.20	102.7%

percentage savings increase slightly as the coal-fired plant costs increase. When the coal plant cost is estimated at \$45/MW-hr, Case 3 has a cost savings of 5.0% compared to Case 1, compared to the cost savings of only 3.9% when the coal plant cost is estimated at \$25/MW-hr. The cost savings increases on a percentage basis as the cost of the coal-fired plant increases because the increased power output associated with Cases 2-4 is worth more under these conditions. Also, Table 6-2 demonstrates that, in all three coal plant cost scenarios, the effect of adding costs for CO_2 recovery as well as decreasing the net power plant output results in an approximate doubling of the cost of electricity when CO_2 recovery is added to a coal-fired power plant.

6.2 Cost of CO_2 Avoidance

Table 6-3 illustrates the cost of CO_2 avoidance for the four cases. As shown in the table, the base cost of CO_2 avoidance for Case 1 is \$44.89/tonne CO_2. The integration of heat recovery in Case 2 achieves a 4.6% reduction in the cost of CO_2 removal, while the addition of the multipressure stripper in Case 3 creates a cost savings of 9.8% over Case 1. Case 4, which includes the multipressure stripper without the heat recovery, leads to a cost savings of 8.4%.

Table 6-4 documents the sensitivity of CO_2 avoidance costs and the percentage reductions in costs as they related to the coal-fired power plant costs. As expected, the absolute costs increase as the coal plant costs increase because the derating caused by parasitic power consumption now has a higher dollar value. For all three coal plant costs (\$25/MW-hr, \$35/MW-hr, and \$45/MW-hr) considered in this analysis, the cost savings is the greatest for Case 3 and is typically around 10%.

Table 6-3. Summary of Cost of CO_2 Avoidance for a Gross 500 MW Coal-Fired Power Plant

	Case 1	Case 2	Case 3	Case 4
Net Power Plant Output after Derating (Base Plant[1] and CO2 Capture and Compression) , MW	280	293	297	293
Total Reduction in Net Power Rating of 453 MW due to Parasitic Loads	38.1%	35.2%	34.4%	35.3%
Base cost of electricity (prior to installation of CO2 Removal), $/MWh	$25.00	$25.00	$25.00	$25.00
Cost of electricity after installation of CO2 Removal, $/MWh	$63.24	$61.81	$59.88	$60.32
Base emissions (without CO2 Capture)				
Tonnes/year (based on 85 % capacity factor)	3.43E+06	3.43E+06	3.43E+06	3.43E+06
Tonnes/MWh	1.016130012	1.016130012	1.016130012	1.016130012
CO2 Emissions with CO2 Capture, based on 90 % removal in absorber				
Tonnes/year	3.43E+05	3.43E+05	3.43E+05	3.43E+05
Tonnes/MWh	0.164214522	0.156839028	0.154887513	0.157085625
Cost of CO2 Avoidance, $/tonne	$44.89	$42.83	$40.50	$41.12
% reduction from Case 1	--	4.6	9.8	8.4

Notes:

1) Base plant includes electricity for PC, ESP, FGD, and SCR systems for 500 MW unit (500 MW Gross, 453 MW Net)

Table 6-4. Sensitivity of CO_2 Avoidance Costs to Coal Plant Cost

	Case 1	Case 2		Case 3		Case 4	
Coal Plant Cost	**$/tonne CO2**	**$/tonne CO2**	**% Reduction**	**$/tonne CO2**	**% Reduction**	**$/tonne CO2**	**% Reduction**
$25/MW-hr	44.89	42.83	-4.6%	40.50	-9.8%	41.12	-8.4%
$35/MW-hr	52.09	49.13	-5.7%	46.56	-10.6%	47.45	-8.9%
$45/MW-hr	59.30	55.43	-6.5%	52.62	-11.3%	53.78	-9.3%

6.3 Evaluation of Selective Operation

As described previously, the strategy of selective operation of the amine system (with its large power consumption) involves operating at higher than 90% reduction (e.g., 95+% CO_2 capture) during periods when power demand is lower, and then shutting down the amine system and maintaining it on hot standby for some fraction of time (e.g., 5%) during peak demand periods when the power demand is highest. As a result, this enables the plant to achieve an overall CO_2 recovery of ~90% on an annualized basis, while minimizing the installation and operation of potentially more expensive peak generation capacity.

To evaluate this option, Cases 1 and 2 were re-run to get 95% removal, with the same costing methodology applied. Table 6-5 presents a comparison of the cost of electricity for these different cases based on continuous operation at design removal for each case (i.e., no peaking has been factored into Table 6-5). As shown in Table 6-5, achieving an extra 5% CO_2 removal increases the cost of Case 1b by approximately 11.9% over Case 1. This is driven solely by the increased capital and operating costs of the MEA system. A comparison of Cases 2 and 2b demonstrates the same trend, with Case 2b having a cost of electricity estimated to be 10.4% higher than Case 2.

Performing the peaking comparison involved making a number of assumptions. For this analysis, the Case 1 and 2 systems were operated continuously for all 7,446 hours of operation allowed by the constraint of an 85% capacity factor. By contrast, Cases 1b and 2b were operated at the 95% removal rate for 7,074 hours/year (or 95% of the operating time), with the CO_2 removal system was turned off during the other 372 hours.

During the “base” period operation of 7,074 hours/year, Cases 1 and 2 obviously achieve lower CO_2 removal, but these cases have less parasitic power consumption by 21 and 16 MW, respectively, when compared to Cases 1b and 2b. However, during the “peaking” period of operation when the CO_2 removal is turned off or placed on standby

Table 6-5. Overview of Costs and COE for 90 and 95% Cases

	No CO2 removal	Case 1, 90% removal	Case 1b, 95% removal	Case 2, 90% removal	Case 2b, 95% removal
Power Plant size, MW	500	500	500	500	500
Net Power Production	453	281	260	294	278
CO2 Recovery, tonne/hr		415	438	415	438
Power Plant Cost, $/MWh	25				
Annual Power Plant Cost	$84,325,950	$84,325,950	$84,325,950	$84,325,950	$84,325,950
CO2 Removal Plant Variable O&M Costs		$7,649,000	$8,527,000	$6,972,000	$7,695,000
CO2 Removal Plant Fixed O&M Costs		$5,247,000	$5,725,000	$5,729,000	$6,359,000
CO2 Removal Plant Capital Recovery Costs		$34,961,000	$38,219,000	$38,224,000	$42,505,000
Total CO2 Removal Costs		$47,857,000	$52,471,000	$50,925,000	$56,559,000
Total Power Plant + CO2 Removal Costs		$132,182,950	$136,796,950	$135,250,950	$140,884,950
Electricity Costs, $/MW-hr produced		63.24	70.75	61.81	68.15
% Increase in COE (vs. Case 1)			11.9%	-2.3%	7.8%
CO2 Emitted, tonne/hr		46	23	46	23

for Cases 1b and 2b, these cases produce a full 453 MW of power due to the 95% capture rate during off-peak demand, or 172 and 159 MW more than Cases 1 and 2.

Table 6-6 presents the cost comparison for these cases. To assess the costs of the effects of the varying power production during peaking and non-peaking periods, the plants in Cases 1 and 2 had to buy peaking power at $130/MW-hr during the 372 hours that Cases 1b and 2b were turned off and selling their full 453 MW to the grid; the $130/MW-hr was chosen with utility guidance and is based on natural gas-fired peaking turbines operating in the more recent high gas price environment seen during 2004. However, Cases 1b and 2b had to buy baseload power to make up their 21 and 16 MW shortfalls during the other 7,074 hours of operation, and it was assumed that they were buying baseload power that had CO_2 controls applied. Based on the costs of electricity shown in Table 6-5, a cost of $70/MW-hr was assigned to this electricity supplementation. This additional electricity purchase assigned to each case meant that all four cases produced the same amount of electricity.

As shown in Table 6-6, the costs of Cases 1b and 2b were higher than the costs of Cases 1 and 2, meaning that the strategy of selective operation was not viable for the assumptions included in this analysis. Case 1b was 4.5% higher for the COE than Case 1a, and Case 2b was 4.0% higher. This represents a narrowing of the gap observed in Table 6-5, but there was not enough cost savings provided by avoiding the higher cost peaking power to overcome the derating caused by achieving 95% removal.

Fundamentally, the analysis of the selective operation approach becomes a tradeoff between the value of the derated power (21 and 16 MW in these two cases) at baseload conditions for 95% of the year and the value of the peaking power (172 and 159 MW in these examples) for 5% of the year. For the base load cost of $70/MW-hr, the

Table 6-6. Evaluation of Selective Operation of 95% Removal System Compared to Continuous Operation of 90% Removal System

		Case 1	Case 1b	Case 2	Case 2b
Base Electricity Rate, MW		281	260	294	278
Base Electricity, hrs		7074	7074	7074	7074
Peak Electricity Rate, MW		453	453	453	453
Peak Electricity, hrs		372	372	372	372
Peaking Electricity Added, MW		172	193	159	175
Peaking Electricity Cost, $/MW		130		130	
Price of Baseload Supplementation from Grid, $/MW			70		70
Annual Power Plant Cost @$25/MW-hr		$84,325,950	$84,325,950	$84,325,950	$84,325,950
CO2 Removal Plant Variable O&M Costs		$7,649,000	$8,100,650	$6,972,000	$7,310,250
CO2 Removal Plant Fixed O&M Costs		$5,247,000	$5,725,000	$5,729,000	$6,359,000
CO2 Removal Plant Capital Recovery Costs		$34,961,000	$38,219,000	$38,224,000	$42,505,000
Total CO2 Removal Costs		$47,857,000	$52,044,650	$50,925,000	$56,174,250
Total Power Plant + CO2 Removal Costs		$132,182,950	$136,370,600	$135,250,950	$140,500,200
Total Peaking Power Costs		$8,338,747		$7,700,764	
Total Grid Supplementation			$10,414,537		$8,048,377
TOTAL POWER & CO2 COSTS		$140,521,697	$146,785,137	$142,951,714	$148,548,577
TOTAL ELECTRICITY, MW-HRS		2,154,298	2,154,298	2,247,542	2,247,542
Cost, $/MW-hr		$65.23	$68.14	$63.60	$66.09
% Increase COE vs. Case 1			4.5%	-2.5%	1.3%

peaking power cost would have to be \$230-240/MW-hr for the cases to have approximately the same cost of electricity. Alternately, if the peaking power cost is maintained at \$130/MW-hr, then the cost of buying power from the grid would have to be around \$25/MW-hr for these cases to have roughly the same cost of electricity.

One issue not evaluated in this analysis of selective operation was the cost or value of CO_2 emissions that were not controlled, which is different in both the 90% removal case and the 95% removal cases. The 90% cases were allowed to buy peaking power from natural gas-fired turbines that did not have CO_2 controls applied to them, while the 95% cases bought power from base load plants that did have CO_2 controls applied. This seems to be a reasonable future regulatory scenario, since controlling large, continuously running base load plants would be much more cost-effective than controlling smaller turbines that run relatively few hours per year. However, it does provide some advantage to the 90% case, since it is able to make up its peaking shortfall with electricity that is produced without the cost of CO_2 control.

6.4 Sensitivity to Capacity Factor

The effect of capacity factor was evaluated to determine the effect on the cost of CO_2 avoidance and overall reduction from the base case. The costs in Table 6-1 were recalculated assuming a 75% capacity factor instead of the 85% used in the study. In this analysis, it was assumed that \$10/MWh of the total \$25/MWh power plant cost (at 85% capacity) was associated with capital and would remain fixed. The remaining \$15/MWh was adjusted based on a 75% capacity factor so that the total power plant cost is now \$27/MWh.

Table 6-7 shows how the capacity factor impacts the results of the study. As shown in the table, the cost of CO_2 avoidance is more sensitive to the capacity factor; however, the overall % reduction from the base case does not change significantly.

Table 6-7. Sensitivity to Capacity Factor (75% versus 85%)

Parameter	Case 1	Case 2	Case 3	Case 4
Cost of CO_2 avoidance, $/tonne (85% capacity factor)	44.89	42.83	40.5	41.12
Cost of CO_2 avoidance, $/tonne (75% capacity factor)	51.68	49.53	46.91	47.56
% reduction from Case 1 (85% capacity factor)	--	4.6	9.8	8.4
% reduction from Case 1 (75% capacity factor)	--	4.2	9.2	8

7.0 SUMMARY AND CONCLUSIONS

This section summarizes the work completed under this Phase I SBIR project and presents the major findings and conclusions, along with an estimate of the technical and economic feasibility.

This project furthers the previous work done by the University of Texas at Austin that proposed process schemes that could theoretically reduce energy costs by 5% to 20% for capturing CO_2 from coal-fired power plants using MEA. The overall objective of this project was to identify additional ways to reduce costs as well as to determine the optimal approach for implementing these energy saving ideas at acceptable capital costs. The specific objectives of this Phase I project were to:

- Develop process designs for approximately three innovative MEA stripper configurations to reduce parasitic energy requirements for CO_2 capture with MEA;
- Develop and evaluate other novel processing schemes that are discovered as a result of process design, engineering, or integration planning;
- Evaluate equipment options and select equipment with the best combination of operability and economics to implement the process designs; and
- Determine how to best integrate the MEA process and CO_2 compression into a coal-fired utility so as to accomplish 90% CO_2 removal at least cost.

The work plan for achieving these objectives consisted of four main tasks. The first task was for process design and simulation in which the basic designs and material/energy balances were developed. The second task was to select and size the equipment for full-scale CO_2 capture systems based on the results of the process design task. The third task was for operations design, in which we worked with utility staff at the Platte River Authority to develop optimal approaches for integrating a CO_2 capture project into a utility. The fourth task was to prepare capital/operating costs and develop economics for each process option evaluated.

Four process configurations and six cases were evaluated (as described in detail in previous sections):

- Case 1: The base case (conventional MEA absorption/regeneration)
- Case 2: Vapor recompression heat recovery
- Case 3: Multipressure stripping with vapor recompression heat recovery
- Case 4: Multipressure stripping without vapor recompression heat recovery

The design basis for these evaluations was a 500 MW gross conventional coal-fired power plant. The CO_2 capture system was based on a generic 30 wt % MEA absorption/regeneration process. The flue gas composition was based on previous wok done at the University of Texas, and the coal composition and heating value were based on DOE guidelines for Illinois No. 6 Coal.

All four process configurations were evaluated based on 90 % removal of CO_2 in the absorber. For Case 1 and 2, an additional case was run (referred to as Case 1a and Case 2a) in which the CO_2 removal specification was 95 %. These latter two cases were run for the purpose of evaluating the feasibility of integrating the CO_2 capture operations with power plant peaking to reduce costs. Under this scenario, the CO_2 capture and compression equipment would be operated selectively at 95 % capture efficiency during non-peaking periods and then the system would be maintained on hot standby for roughly 5 % of the time during peak demand periods when the electricity is most valuable. As a result, the plant would achieve an overall 90 % capture efficiency on an annualized basis using this “95/5” peaking approach.

The major conclusions of this work are summarized in the following paragraphs:

- Reductions in the cost of CO_2 capture (\$/tonne CO_2 avoided) ranged from 4.6 to 9.8 percent among the cases;

- The configuration with the least cost per tonne avoided was Case 3 (multipressure stripping with vapor recompression);

- The parasitic energy load (as defined by the difference in net power production before and after CO_2 capture/compression equipment is installed) could be reduced by 7.5 – 9.8 percent, freeing up 13 – 17 MW of power for sale to the grid based on the model 500 MW (gross) power plant used in this study;

- The value of this incremental increase in net power production results in a short payback on capital, approximately six months to one year for Cases 3 and 4 (assuming a value of 0.06 $/kWh), suggesting that these heat integration processes are very likely to be implemented at future CO_2 capture facilities using MEA;

- Reboiler steam requirements were reduced by 18 to 39 percent, which is desirable from the utility operating perspective despite the partially offsetting increases in electrical requirements for the compression train; and

- The 95/5 peaking strategy becomes attractive economically when the value of peak electricity is in excess of approximately $230/MWh. The primary drawback of the 95/5 approach is the increased capital and operating costs and consequent de-rating of the power plant that is required to operate the CO_2 capture equipment 95% of the time at 95% removal vs. 90% removal annually.