# Advanced Amine Solvent Formulations and Process Integration for Near-Term $CO_2$ Capture Success

Final Report

Work Performed Under Grant No.: DE-FG02-06ER84625
Submitted June 28, 2007

to

U.S. Department of Energy
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## DISCLAIMER

## ACKNOWLEDGEMENTS

This report is sponsored by the U.S. Department of Energy's National Energy Technology Center (DOE/NETL) under Contract No. DE-FG02-06ER84625. The authors would like to express sincere appreciation for the support and guidance of the DOE/NETL project manager, Jose D. Figueroa. The authors would also like to thank Luminant for its support and advice throughout the project.

# ABSTRACT

This Phase I SBIR project investigated the economic and technical feasibility of advanced amine scrubbing systems for post-combustion $CO_2$ capture at coal-fired power plants. Numerous combinations of advanced solvent formulations and process configurations were screened for energy requirements, and three cases were selected for detailed analysis: a monoethanolamine (MEA) base case and two "advanced" cases: an MEA/Piperazine (PZ) case, and a methyldiethanolamine (MDEA) / PZ case. The MEA/PZ and MDEA/PZ cases employed an advanced "double matrix" stripper configuration. The basis for calculations was a model plant with a gross capacity of 500 MWe. Results indicated that $CO_2$ capture increased the base cost of electricity from 5 cents/kWh to 10.7 c/kWh for the MEA base case, 10.1 c/kWh for the MEA / PZ double matrix, and 9.7 c/kWh for the MDEA / PZ double matrix. The corresponding cost per metric tonne $CO_2$ avoided was 67.20 \$/tonne $CO_{2,}$ 60.19 \$/tonne $CO_{2,}$ and 55.05 \$/tonne $CO_{2,}$ respectively. Derated capacities, including base plant auxiliary load of 29 MWe, were 339 MWe for the base case, 356 MWe for the MEA/PZ double matrix, and 378 MWe for the MDEA / PZ double matrix. When compared to the base case, systems employing advanced solvent formulations and process configurations were estimated to reduce reboiler steam requirements by 20 to 44%, to reduce derating due to $CO_2$ capture by 13 to 30%, and to reduce the cost of $CO_2$ avoided by 10 to 18%. These results demonstrate the potential for significant improvements in the overall economics of $CO_2$ capture via advanced solvent formulations and process configurations.

# TABLE OF CONTENTS

## TABLE OF CONTENTS (CONTINUED)

## LIST OF FIGURES

# LIST TABLES

# EXECUTIVE SUMMARY


This Phase I SBIR project investigated the economic and technical feasibility of advanced amine scrubbing systems for post-combustion $CO_2$ capture at coal-fired power plants. Amine-based scrubbing is one of the most likely near-term options for post-combustion $CO_2$ capture. Conventional amine scrubbing with monoethanolamine (MEA) and simple absorption and stripping flow configurations can achieve 90% $CO_2$ capture. However, the capital and operating costs are very high; work conducted under a previous DOE SBIR grant (DE-FG02-04ER84111) estimated that amine-based $CO_2$ capture would increase the cost of electricity by 3.8 cents/kWh in 2004 dollars and material costs. Therefore, this project investigated systems employing advanced amine solvent formulations and process configurations in order to reduce capital and operating costs. Trimeric Corporation completed this project with a subcontract to the University of Texas and with in-kind assistance from the Dow Gas Treating Services Group and Luminant.

First, the energy requirements for a large array of solvents and process configurations were evaluated in a screening study. Then, three cases were selected for detailed, rigorous analysis: one base case and two "advanced" cases, which employed methyldiethanolamine (MDEA) and piperazine (PZ).

**Cases Selected for Detailed Analysis**

| Case Name | Solvent | Configuration |
|---|---|---|
| Base Case | 7 m MEA | Conventional |
| MEA / PZ double matrix | 7 m MEA, 2 m PZ | Double Matrix |
| MDEA / PZ double matrix | Proprietary concentrations | Double Matrix |

*Note: "m" equals molal.*

Next, rigorous process simulations with mass and energy balances were prepared. Then, equipment was sized and selected, and purchased equipment costs were developed. Finally, capital costs, operating costs, incremental cost of electricity, and cost of avoided $CO_2$ emissions were estimated.

The design basis for these evaluations was a 500 MW gross conventional coal-fired power plant using Illinois #6 subbituminous coal. A wet flue gas desulfurization (FGD) unit was assumed to be located upstream of the $CO_2$ capture unit. The target $CO_2$ removal was 90%. Any captured $CO_2$ was delivered at pipeline pressure (15.2 MPa, 2200 psia). The entire $CO_2$ capture systems consisted of a single inlet gas train, multiple parallel amine units, and a single, common $CO_2$ compression train.

Results estimated that $CO_2$ capture increased the base cost of electricity from 5 cents/kWh to 10.7 c/kWh for the MEA base case, 10.1 c/kWh for the MEA / PZ double matrix, and 9.7 c/kWh for the MDEA / PZ double matrix. The corresponding cost per metric tonne $CO_2$

avoided was 67.20 \$/tonne $CO_2$, 60.19 \$/tonne $CO_2$, and 55.05 \$/tonne $CO_2$, respectively. Derated capacities, including base plant auxiliary load of 29 MWe, were 339 MWe for the base case, 356 MWe for the MEA/PZ double matrix, and 378 MWe for the MDEA / PZ double matrix. When compared to the base case, systems employing advanced solvent formulations and process configurations were estimated to reduce reboiler steam requirements by 20 to 44%, to reduce derating due to $CO_2$ capture by 13 to 30%, and to reduce the cost of $CO_2$ avoided by 10 to 18%. These results, summarized in the table below, demonstrate the potential for significant improvements in the overall economics of $CO_2$ capture via advanced solvent formulations and process configurations.

**Summary Results of Derating, Cost of Electricity, and Cost of $CO_2$ Avoided**

| Description | Units | MEA Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| Gross generating capacity | MWe | 500 | 500 | 500 |
| Net generating capacity without $CO_2$ capture | MWe | 471 | 471 | 471 |
| Net generating capacity with $CO_2$ capture | MWe | 339 | 356 | 378 |
| Derating due to $CO_2$ capture | MWe | 132 | 115 | 93 |
| Reduction in derating due to $CO_2$ capture | % | | 13 | 30 |
| Base plant cost of electricity | c/kWh | 5.0 | 5.0 | 5.0 |
| Total COE | c/kWh | 10.7 | 10.1 | 9.7 |
| Increase in COE | % | 113 | 102 | 95 |
| Cost of $CO_2$ avoided | \$/tonne | 67.20 | 60.19 | 55.05 |
| Reduction in cost of $CO_2$ avoided | % | | 10 | 18 |

# 1.0 INTRODUCTION

This report documents the methodology and results of Trimeric Corporation's Small Business Innovative Research (SBIR) Phase I project, "Advanced Amine Solvent Formulations and Process Integration for Near-Term $CO_2$ Capture Success" (DOE Grant No. DE-FG02-06ER84625). 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

The United States has vast reserves of coal. These abundant resources will play a key role in meeting our country's near-term energy demand while maintaining economic security. However, the use of coal in conventional coal-fired power plants emits large quantities of the greenhouse gas (GHG) carbon dioxide ($CO_2$) . Climate change science suggests that higher atmospheric GHG concentrations may cause changes in the global climate. Since the consequences of changes in global climate are potentially very significant, there is strong interest in reducing the amount of anthropogenic $CO_2$ emissions. As a result of these concerns, the U.S. Department of Energy (DOE) National Energy Technology Laboratory (NETL) is supporting the development of technologies that improve the environmental soundness and economic viability of fossil fuel extraction and use.

To address global warming concerns, President Bush 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% by 2012 (NETL, 2007). $CO_2$ emissions from electric power production contributes about 33% of U.S. GHG emissions (DOE May 2005); any effort to reduce greenhouse gas intensity virtually must address this sector. Therefore, the DOE's NETL is supporting the development of technologies that capture and subsequently sequester $CO_2$ from coal-fired power plants. Specifically, the DOE's goal is to achieve 90% $CO_2$ capture with 99% storage permanence at less than a 20% increase in the cost of energy services by 2012 (DOE May 2005).

$CO_2$ capture technologies are divided into three broad categories: post-combustion, pre-combustion, and oxy-fuel. Of these, post combustion capture may be the most challenging, because the flue gas is at a low pressure and the $CO_2$ is dilute, which makes $CO_2$ capture more difficult and increases sequestration compression costs. However, post-combustion technology is the only category that applies to over 98% of existing fossil power production assets. Thus, in order to meet the President's goal of 18% reduction in GHG intensity by 2012, a key practical target is a post-combustion technology that achieves the DOE's performance and cost goals. The technology research conducted under this contract addresses this post-combustion category and works toward the achievement of the DOE's goals.

Amine-based scrubbing is one of the most likely near-term options for post-combustion $CO_2$ capture. Conventional amine scrubbing can achieve 90% $CO_2$ capture; however, the capital and operating costs are very high. In a FY2005 SBIR project, Integrating MEA Regeneration with $CO_2$ Compression and Peaking to Reduce $CO_2$ Capture Costs (DE-FG02-04ER84111), Trimeric and the University of Texas (UT) demonstrated that using heat integration and alternate process configurations can decrease overall monoethanolamine (MEA) scrubbing costs by nearly 10% (Fisher, 2005). While this was encouraging, further reductions in capital and operating costs are required to meet the DOE performance goals.

The economic analysis from the previous SBIR project indicates what areas to target for capital and operating cost savings. The operating costs dominate the overall capture costs because $CO_2$ capture with a conventional MEA system derates a 500 MWe gross capacity plant by an additional 173 MWe beyond the base plant auxiliary loads, which corresponds to a gross capacity derating of more than one third (Fisher, 2005). For a conventional MEA system, energy requirements of the stripper reboiler and the compressor account for nearly 90% of the derating. Process configurations that will have the greatest impact on cost focus on lowering stripper reboiler and compressor energy costs. Details of the analysis carried out under the previous SBIR grant may be found in the final report for that project (Fisher 2005).

In addition to the current high cost of amine treatment, some operational challenges impede the adoption of flue gas amine scrubbing. First, residual oxygen, sulfur dioxide ($SO_2$), and other species in the flue gas can chemically degrade the amine (Goff, 2005). Process heat also thermally degrades amine solvents over time. Second, the amine liquid solution can corrode process equipment and often requires corrosion inhibitors. Alternate amine solvents can avoid these problems; for example, solvents formulated with piperazine do not undergo the same thermal and chemical degradation mechanisms as the conventional monoethanolamine solvent.

### 1.2 Research Objectives

This project studied improved amine-based $CO_2$ capture system, where a system comprises a solvent and a process configuration. These systems sought to reduce stripper reboiler energy costs and reduce solvent degradation costs. Specifically, the research objectives for this SBIR project included the following:

- Establish the two most promising systems of solvent formulation and process scheme based on a screening of several systems;
- Estimate the capital and operating costs of these top two systems;
- Compare these economics with an “updated” baseline MEA configuration and with DOE targets;
- Resolve how the amine and the compression systems will integrate with the power plant; and
- Select the best process configuration and solvent formulation for future pilot testing.

These technical objectives in Phase I lay the groundwork for continued commercialization efforts. In addition, the current research leverages extensive laboratory work already conducted or scheduled for completion by a research group at the University of Texas led by Dr. Gary T. Rochelle.

### 1.3 Project Participants

Trimeric Corporation (Trimeric) served as the prime contractor for this project. The University of Texas (UT) subcontracted to Trimeric. Dr. Gary Rochelle of the University of Texas and his research group performed the process simulations and provided general technical insight and guidance. Dr. Craig Schubert of the Dow Gas Treating Services group provided input on industrial solvents and provided process simulations. Luminant 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, the screening study, and the cases selected for detailed analysis;
- 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 presents the costs of the three detailed cases in terms of the DOE NETL metrics, cost of electricity and cost of avoided $CO_2$ emissions; and
- Section 7: Summary and Conclusions presents the findings of the research.

**References (Section 1)**

DOE NETL. “Carbon Sequestration Technology Roadmap and Program Plan – 2005”, May 2005.

Fisher, K. S., C. M. Beitler, C. O. Rueter, K. Searcy, G. T. Rochelle, and M. Jassim. “Integrating MEA Regeneration with CO2 Compression and Peaking to Reduce $CO_2$ Capture Costs.” Final Report under DOE Grant DE-FG02-04ER84111, June 9, 2005.

Goff, G.S., Oxidative Degradation of Monoethanolamine in $CO_2$ Capture Processes: Iron and Copper Catalysis, Inhibition, and $O_2$ Mass Transfer," PhD Dissertation, The University of Texas at Austin, 2005.

NETL 2007. (http://www.netl.doe.gov/technologies/carbon_seq/refshelf/fact_sheets/HowWeFit_global.pdf)

## 2.0 CONCEPTUAL APPROACH

This section discusses the conceptual approach that was used on the project. Early in the project, the UT group performed a screening study to evaluate energy requirements for a large array of solvents and process configurations. Then, the project team selected three cases for detailed, rigorous analysis: one base case and two “advanced” cases. Next, team members prepared rigorous process simulations with mass and energy balances. Trimeric then sized and selected equipment and developed purchased equipment costs. Finally, Trimeric calculated capital costs, operating costs, incremental cost of electricity, and cost of avoided $CO_2$ emissions.

The following sections provide more detail on the solvents and configurations considered in the screening study, the design basis, and the engineering and economic analysis approach.

### 2.1 Improved Solvents and Process Configurations

#### 2.1.1 Solvents

Monoethanolamine (MEA) is the conventional amine solvent selected for $CO_2$ scrubbing. However, MEA has several disadvantages when treating flue gas: chemical degradation, thermal degradation, and corrosivity. The UT research group led by Dr. Gary Rochelle has studied several improved solvent formulations that seek to overcome the obstacles associated with conventional MEA. A solvent formulation refers to a mixture of solvents with specific concentrations for each component. The important alternative solvents include piperazine-promoted potassium carbonate ($K_2CO_3$) or “KPIP” solvents, piperazine-promoted MEA or MEA/PZ solvents, promoted tertiary amines including piperazine-promoted methyl diethanolamine (MDEA / PZ), and mildly hindered amines. Many of the solvent formulations researched by Dr. Rochelle’s group include solvent components initially developed by the Dow Gas Treating Services Group, which has decades of experience developing alkanolamine solvents, marketing these solvents, and providing technical services to clients with gas treating facilities. As part of Dow’s ongoing research into amine scrubbing, the company has developed the potassium carbonate/piperazine (KPIP) solvents. Under research programs funded by the

DOE (DE-FC26-02NT41440), Dr. Rochelle's group has performed extensive laboratory, bench, and pilot testing with the Dow solvents.

Monoethanolamine promoted piperazine (MEA / PZ) should provide faster $CO_2$ absorption rates and greater capacity for $CO_2$. Piperazine is less prone to thermal degradation than MEA (Rochelle, 2007). Therefore, the capacity of 7 m MEA can be increased substantially by adding 2 m PZ. Because piperazine is currently more expensive than MEA, management of thermal and oxidative degradation of the MEA/PZ solvent formulation will contribute to lower operating costs.

MDEA promoted piperazine (MDEA / PZ) has been used commercially for a number years to remove $CO_2$ from natural gas and hydrogen at $CO_2$ partial pressures greater than those of flue gas. As a tertiary amine, MDEA has the potential for greater $CO_2$ capacity than MEA. The addition of piperazine significantly improves the rate of $CO_2$ absorption. This solvent has not been used with a high-oxygen concentration, so oxidative degradation may be a major concern.

The KPIP solvents have three main differences from MEA: lower heat of $CO_2$ desorption, faster rates of $CO_2$ absorption, and thermal resiliency. The lower heat of $CO_2$ desorption can decrease the reboiler steam requirements. The faster absorption kinetics can create richer solutions given the same absorber capital costs. Thermal resiliency means that the KPIP solvents can operate at higher temperature and pressure without degrading at the same rate MEA would under similar conditions. Thus, KPIP solvents may be more suited to process configurations such as multipressure stripping. However, optimum solvent formulation is yet to be determined and is one piece of the optimization puzzle.

Through DOE cooperative agreement DE-FC26-02NT41440, The UT research group is developing rigorous process models, verified by pilot testing, for absorption and stripping of $CO_2$ with the KPIP solvents. The group is conducting pilot tests to verify the models. The UT contract will also investigate solvent losses, solvent reclamation, and corrosivity. Solvent degradation and reclamation studies will indicate how the KPIP solvents are affected by other contaminants such as $SO_2$, HCl, $NO_x$, etc. Through research to date, the UT team has prioritized the list of promising solvent formulations to two mixtures with varying concentration of piperazine (PZ) and potassium carbonate ($K_2CO_3$): 5 m K+ / 2.5 m PZ ("KPIP5"), and 6.4 m K+ / 1.6 m PZ ("KPIP6.4"). In these formulations, "m" signifies molal concentration.

### 2.1.2 Process Configurations

Several process configurations were considered in this research: conventional, vacuum stripping, double matrix stripping ("double matrix" or "DM"), double matrix vacuum with heat recovery ("DMVHR"), multipressure stripping without heat recovery, and multipressure stripping with heat recovery. Section 2.1.2 introduces these configurations with brief, qualitative descriptions. Specific values for operating conditions vary according to solvent selection; these data are provided for specific cases later in the report.

Figure 2-1 shows a simplified process flow diagram (PFD) for the conventional MEA $CO_2$ capture system. Flue gas flows from the FGD scrubber to the $CO_2$ capture system. Inlet flue gas enters the bottom of the absorber. Cool lean amine enters the top of the absorber. The amine absorbs $CO_2$ as it flows downward and contacts the gas. Rich amine exits the bottom of the absorber. Flue gas exits from the top of the absorber and flows to the stack. Rich amine exchanges heat with hot lean amine. The preheated rich amine then flows to the stripper, where $CO_2$ desorbs from the amine solution. Warm stripper overheads flow to the stripper condenser, where the vapor is cooled and water is condensed. The remaining low-pressure $CO_2$ vapor then flows to compression. A stripper reboiler provides heat for the $CO_2$ desorption. Hot lean amine exits from bottom of the stripper and is cooled through cross exchange with the rich amine. Water enters the system at the top of the absorber, where it serves as a water wash section and decreases amine losses with the sweet gas. Steam provides heat to the reboiler.

Vacuum stripping is essentially the same as the conventional configuration except that the stripper is operated at lower, vacuum pressures.

Figure 2-2 shows a simplified PFD for the double matrix configuration. This configuration reduces compression work because a large portion of the $CO_2$ is stripped at higher pressures. The energy required for separations in the double matrix stripper system may more closely approach the ideal energy requirements because the solvent is more isothermal.

Rich amine exits the bottom of the absorber. Downstream of the rich amine pump, the flow splits into two separate streams: one sent to the Low Pressure (LP) Stripper and one sent to the High Pressure (HP) stripper. The split between LP and HP streams is optimized according to the selected pressures and solvent formulation. The "LP" rich amine is preheated via exchange with warm semi-lean amine. LP rich amine enters the top of the LP stripper upper section and contacts gas from the bottom section as the liquid falls through the packed section. Semi-lean amine exits from the bottom of the upper section of the LP Stripper. Warm, semi-lean amine is cooled via exchange with the LP rich amine. The semi-lean amine then flows to the middle of the absorber. Vapors from the upper section of the LP Stripper flow to the LP Condenser. LP Condensate flow to the makeup water system, and remaining vapor flows to the first stage of compression. A rich amine booster pump provides additional driving force to move the rich amine into the HP stripper. The "HP" rich amine is preheated via exchange with hot lean amine. The warm, HP rich amine enters the top of the HP stripper. Vapors exiting the top of the HP stripper combine with gas exiting the first stage of compression. This combined vapor stream flows to the 1$^{st}$ interstage cooler. Condensate from this cooler also returns to the makeup water system, and the remaining vapor flows to the latter stages of compression. The HP reboiler provides heat to desorb CO2 in the HP Stripper. Hot, HP lean amine exits the bottom of the HP Stripper and flows to the top of the bottom LP stripper section, where the liquid flows down through a packed bed and contacts gas generated by the LP Reboiler. Vapor from the bottom section of the LP Stripper flow to the upper section off the LP Stripper. Hot LP lean amine exits from the bottom of the lower section of the LP Stripper and is cooled via exchange with HP rich amine. Then, the LP lean amine flows to the top of the absorber. Steam provides heat for both reboilers.

Figure 2-3 shows the simplified PFD for the Double Matrix Vacuum with Heat Recovery (DMVHR) flow scheme. The configuration is very similar to the double matrix except that the lower pressure stripper is run at vacuum conditions and the outlet gas from the 1st stage of compression provides some reboiler heat before mixing with the HP Stripper overheads. Steam provides the remaining heat for the HP Reboiler and all the heat for the LP Reboiler.

The PFD for multipressure stripping without heat recovery is shown in Figure 2-4. In this case, stripping occurs at several pressure increments. The vapor streams exiting lower-pressure stripper segments are compressed and serve as the entering vapor streams to the next higher pressure stripper segment. The reboiler provides heat for the lowest pressure stripping segment, and the heat of compression provides stripping heat for the higher-pressure segments. In comparison, the matrix stripping configurations have separate reboilers for each stripper and do not compress overhead vapors from lower pressure strippers to provide vapor for higher pressure strippers. In multipressure stripping, the liquid streams exit higher-pressure stripper segments and enter lower-pressure segments. The highest pressure stripper has a reflux condenser. Vapor exits this condenser and flows to the remaining stages of compression, which have water-cooled interstage coolers.

The PFD for multipressure stripping with heat recovery is shown in Figure 2-5. This configuration is similar to the multipressure stripping without heat recover except that the reflux condenser is eliminated and hot outlet gas from the latter compression stages provides some heat for the reboiler. In this configuration, water cooling does supplement any interstage cooling of the $CO_2$ after it provides heat to the reboiler. Thus the interstage cooling temperature is higher in this configuration when compared to configurations with water-cooled interstage cooling.

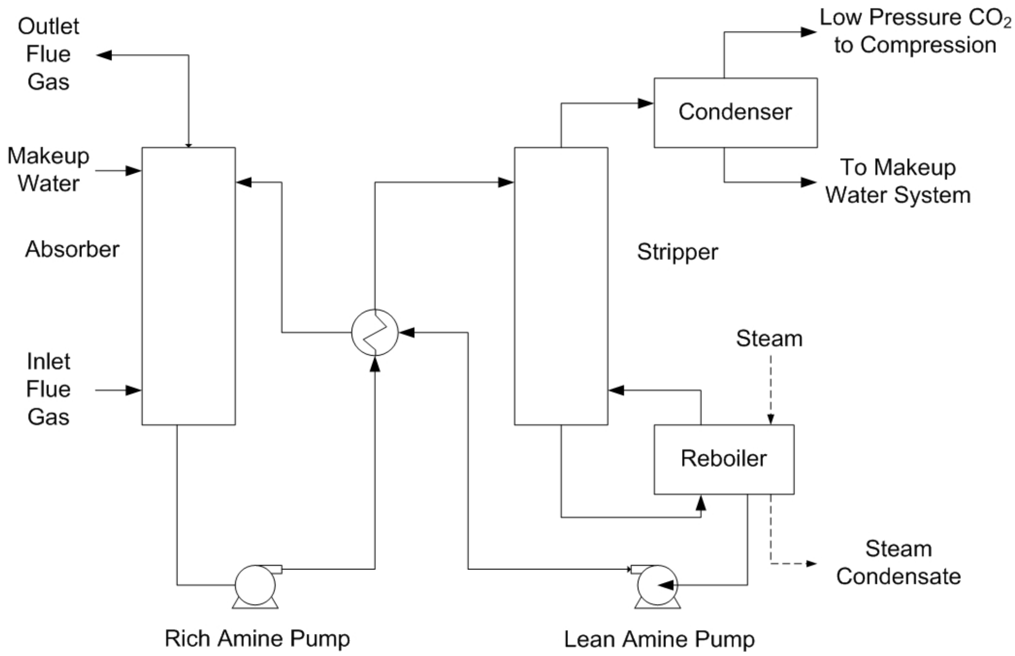


**Figure 2-1. Base Case - Simplified PFD**

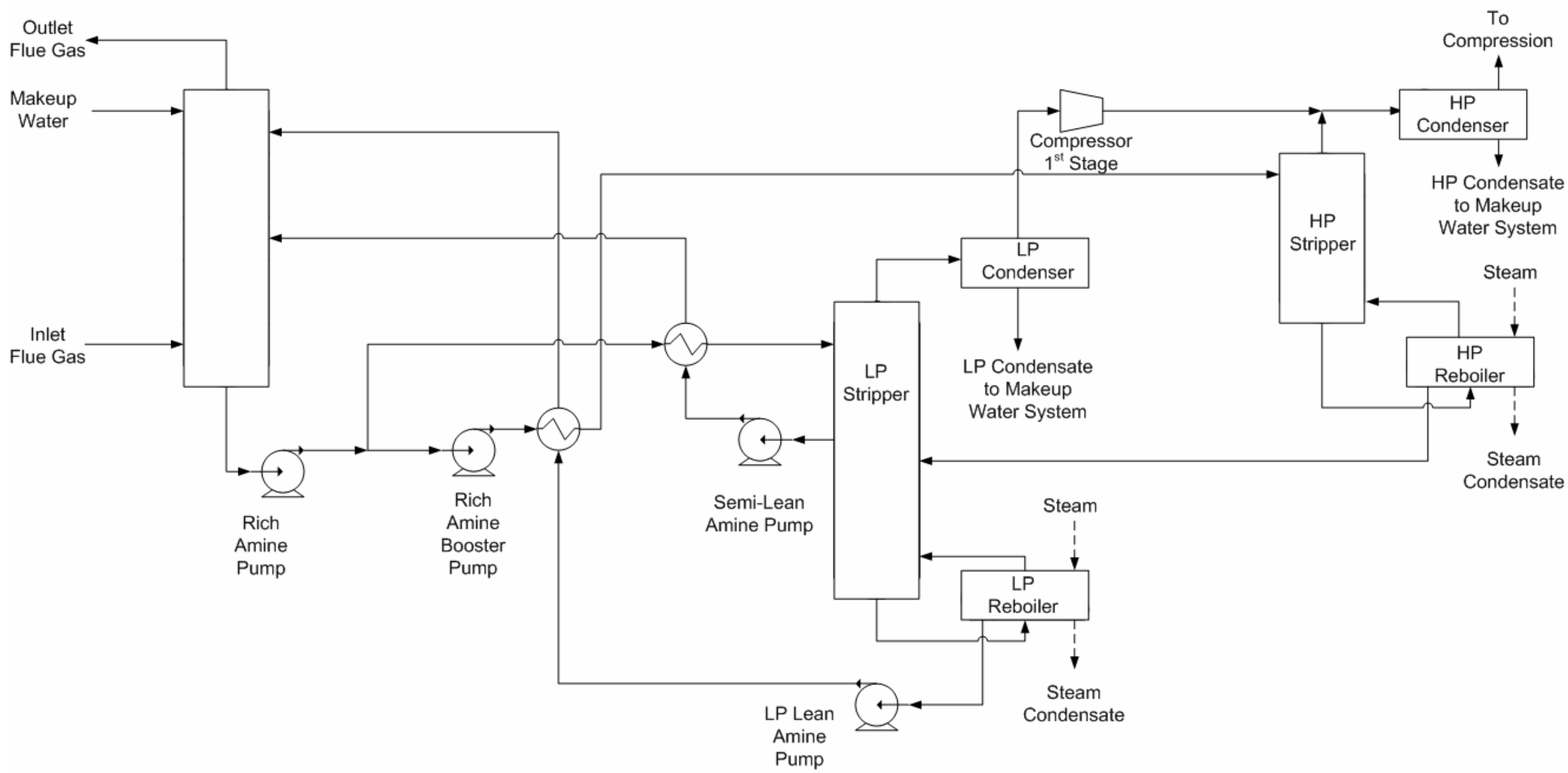


**Figure 2-2. Double Matrix - Simplified PFD**

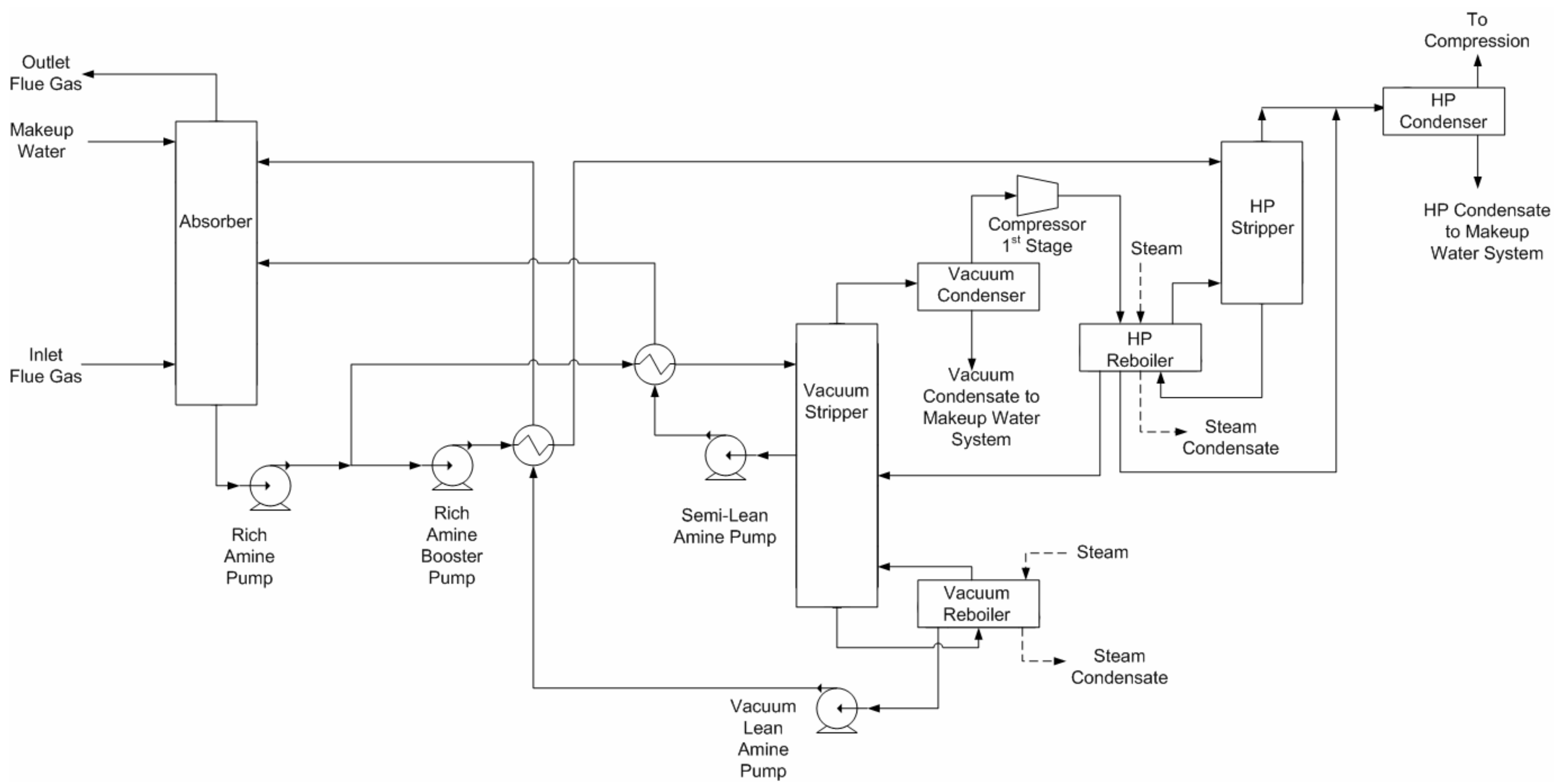


Figure 2-3. Double Matrix Vacuum with Heat Recovery - Simplified PFD

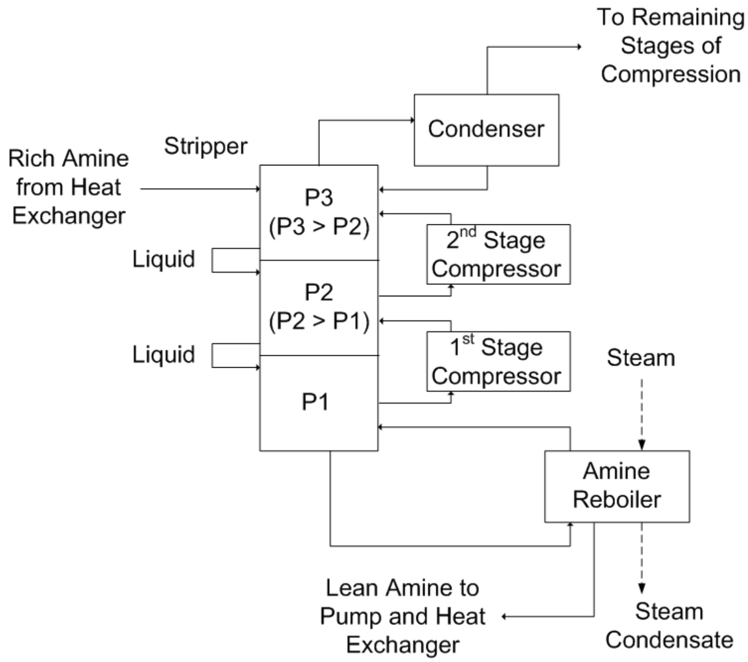


**Figure 2-4. Multipressure Stripping without Heat Recovery – Simplified PFD**

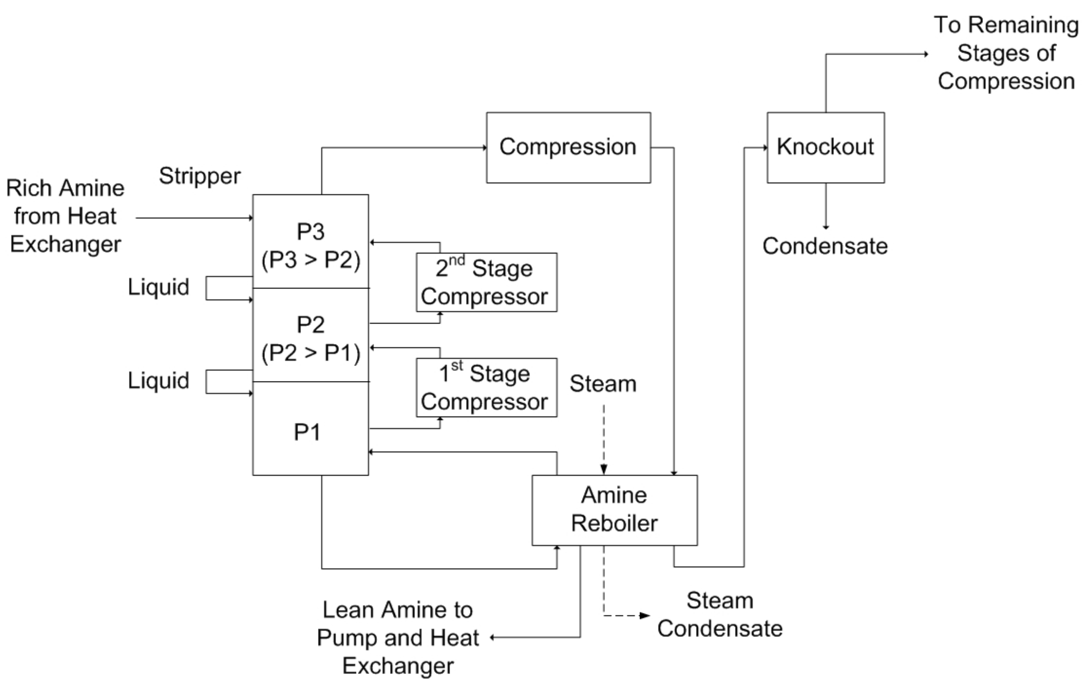


**Figure 2-5. Multipressure Stripping with Heat Recovery – Simplified PFD**

## 2.2 Process Simulation Design Basis

The design basis consists of power plant type, gross generating capacity, fuel, FGD type, $CO_2$ removal, $CO_2$ product specifications, and operating parameters for the $CO_2$ capture system. Table 2-1 shows the design basis.

The primary analyses are based on a 500 MW, pulverized-coal-fired supercritical boiler with a wet FGD system. The unit fires Illinois #6 subbituminous coal, and calculations were based on the ultimate analysis provided in the NETL Systems Analysis Guidelines (DOE 2005). The inlet flue gas composition was calculated using a gross heat rate of 9,674 Btu/kWh and 28% excess air. The resulting flue gas composition, flow rate, and conditions are shown in Table 2-2. Sulfur dioxide is not included in this composition; any remaining sulfur dioxide will be scrubbed with caustic in a direct-contact cooler prior to entering the $CO_2$ capture scrubber.

# Table 2-1. Design Basis – Process Inputs

| Description | SI Units | Value | English Units | Value |
|---|---|---|---|---|
| **General** | | | | |
| Type | - | Pulverized coal-fired supercritical boiler | | |
| Gross Capacity | MWe | 500 | | |
| FGD | - | Yes, wet FGD | | |
| Capacity Factor | % | 80 | | |
| Combustion excess air | % | 28 | | |
| Gross heat rate | - | | Btu/kWh | 9,674 |
| CO2 removal | % | 90 | | |
| **Stream Data** | | | | |
| *Solvent* | | | | |
| Solvents for Detailed Analysis: | | | | |
| Conventional MEA | - | MEA (7 m, ~30 wt%) | | |
| Promoted MEA | - | MEA / PZ (7 m MEA, 2 m PZ) | | |
| MDEA | - | MDEA (~50 wt%) | | |
| Solvents for Screening Study Only: | | | | |
| Potassium Carbonate / Piperazine 1 | | KPIP 4545 (4.5 m K+, 4.5 m PZ) | | |
| Potassium Carbonate / Piperazine 2 | | KPIP 6416 (6.4 m K+, 1.6 m PZ) | | |
| Solvent degradation, leaks, spills | kg/tonne CO2 captured | 1.5 | | |
| *Coal* | | | | |
| General Data | | | | |
| Rank | - | High Volatile Bituminous | | |
| Seam | - | Illinois #6 (Herrin) | | |
| Sample Location | - | St Clair Co., IL | | |
| Ultimate Analyses (wt%) | | | | |
| Moisture | wt% | 7.97 | | |
| Carbon | wt% | 60.42 | | |
| Hydrogen | wt% | 3.89 | | |
| Nitrogen | wt% | 1.07 | | |
| Chlorine | wt% | 0.05 | | |
| Sulfur | wt% | 4.45 | | |
| Ash | wt% | 14.25 | | |
| Oxygen (BD) | wt% | 7.90 | | |
| *Amine Absorber Inlet Flue Gas* | | | | |
| Temperature | C | 40.0 | F | 104.0 |
| Pressure | kPa | 111.67 | psia | 16.20 |
| *Ambient air* | | | | |
| Relative humidity | % | 60 | | |
| Temperature | C | 15 | F | 59 |
| Pressure | kPa | 101.325 | psia | 14.696 |
| Wet bulb temperature | C | 7.2 | F | 45 |
| *Outlet $CO_2$ Specification* | | | | |
| Pressure | bar | 152 | | |
| | kPa | 15200 | psia | 2205 |
| Water content | Dew point (K) | 233 | Dew point (°F) | -40 |
| $N_2$ | ppmv | <300 | | |
| $O_2$ | ppmv | <40 | | |
| Ar | ppmv | <10 | | |
| *Cooling water* | | | | |
| Supply temperature | C | 29.4 | F | 85 |
| Return temperature | C | 43.3 | F | 110 |
| *Steam to Turbine (Compressor Driver)* | | | | |
| *This steam will drive a turbine joined by a common crankshaft to the capture compressors.* | | | | |
| Source of steam | | Intermediate pressure steam | | |
| Temperature | C | 316 | F | 600 |
| Pressure | kPa | 1103 | psia | 160 |
| Superheated? | - | superheated | | |

**Table 2-2. Inlet Flue Gas Conditions**

| | | | SI | | English | |
|---|---|---|---|---|---|---|
| | | | Units | Value | Units | Value |
| | | | | | | |
| *FGD Outlet Flue Gas* | | | | | | |
| | Composition | | | | | |
| | | $N_2$ | mol% | 70.03 | | |
| | | $O_2$ | mol% | 4.65 | | |
| | | $CO_2$ | mol% | 12.38 | | |
| | | $H_2O$ | mol% | 12.94 | | |
| | Flow rate - total | | kmol/s | 23.5463 | | |
| | Flow rate - per train | | kmol/s | 5.887 | lbmol/h | 46,719 |
| | | | sm3/s | 139.4 | MMSCFD | 425.5 |
| | | | dm3/s | 156,706 | acfm | 332,042 |
| | Temperature | | C | 51.2 | F | 124.2 |
| | Pressure | | kPa | 101.325 | psia | 14.696 |
| | | | | | | |

## 2.3 Engineering and Economic Analysis Approach

The following subsections describe in greater detail the screening study, the process simulations, and the engineering and economic analysis approach.

### 2.3.1 Screening Study

Early in the project, the UT group performed a screening study to evaluate energy requirements for a large array of solvent and process configuration combinations. Then, the project team selected three cases for detailed, rigorous analysis: one base case and two "advanced" cases. The base case has been updated with the following changes to the base case used in the previous SBIR (DE-FG02-04ER84111).

- Inlet flue gas is cooled to 40°C (104°F) prior to entering the absorber.
- Rich/lean amine exchanger temperature approach is reduced from 10°C (18°F) to 5°C (9°F).
- Stripper operates at 172 kPa (25 psia) instead of 203 kPa (29.4 psia).

These minor changes were made to ensure that the base case is as representative of an actual MEA-based process design as possible. The flue gas was cooled so that the absorber would operate at a lower temperature and achieve a higher CO2 loading in the rich amine. This higher loading would in turn decrease the circulation rate and reduce energy requirements in the reboiler. The temperature approach in the rich/lean exchanger was decreased because this change was estimated to decrease the total required energy of the unit by ~12% (Oyenekan, 2007). The stripper pressure was decreased slightly to be consistent with low pressure strippers in matrix configurations and thus allow easier comparison between process configurations.

As with the original base case, 7 m (30wt%) MEA was also the solvent in the updated base case. The two "advanced" cases were selected based on the screening study and input from industry advisors as described later in this section. An advanced case uses a novel solvent and process configuration. The following process configurations were considered in the initial screening study:

- Double matrix stripping,
- Vacuum stripping,
- Multipressure stripping without heat recovery,
- Multipresure stripping with heat recovery, and
- Additional combinations thereof.

These configurations were described earlier in Section 2.1.2. The following solvents were included in the screening study:

- MEA (7 m MEA),
- MEA PZ (7 m MEA, 2 m PZ),
- MDEA PZ (proprietary formulation),
- KPIP 4545 (4.5 m K+, 4.5 m PZ), and
- KPIP 6416 (6.4 m K+, 1.6 m PZ).

Total equivalent work was estimated for all screening cases and provided the basis of selection for detailed analysis. The screening study was based largely on material included in the dissertation of Oyenekan while at the University of Texas (Oyenekan, 2007). The double matrix screening cases generally showed the largest decrease in total energy requirements of all configurations, which corresponds to ~20% savings in equivalent work over the previous MEA base case from the 2004 SBIR (DE-FG02-04ER84111) and ~15% over the revised MEA base case. The double matrix configuration was, therefore, selected as one of the configurations that warranted detailed analysis. MEA/PZ and MDEA/PZ decreased total energy requirements by at least 10% when compared to MEA for several different configurations. The best combination of solvent and process configuration was double matrix with MDEA/PZ; thus, this case was selected for detailed study. The concentrations of the MDEA / PZ formulation are proprietary, yet the results from this case will indicate what costs may be anticipated for the range of MDEA / PZ solvent formulations. The systems that were competitive with the MDEA/PZ double matrix were the MEA/PZ double matrix and a KPIP double matrix. MEA/PZ is more developed than the KPIP solvent and has lower perceived risk to industry. Therefore, 7m MEA/ 2m PZ double matrix was selected as the other "advanced" case. To summarize, the cases selected for detailed analysis are shown in Table 2-3.

**Table 2-3. Cases Selected for Detailed Analysis**

| Case Name | Solvent | Configuration |
|---|---|---|
| Base Case | 7 m MEA | Conventional |
| MEA / PZ double matrix | 7 m MEA, 2 m PZ | Double Matrix |
| MDEA / PZ double matrix | Proprietary concentrations | Double Matrix |

### 2.3.2 Process Simulation

UT performed rigorous modeling of the $CO_2$ absorption and stripping for the MEA base case and the MEA/PZ double matrix case. The calculations use AspenONE® and RateSep™ software with advanced calculation methods developed under previous DOE funding (DE-FC26-02NT41440). AspenONE® and RateSep™ are commercial process modeling software supplied by Aspen Technology, Inc. The model accounts for mass transfer with fast reaction in the liquid boundary layer, gas film diffusion, liquid film diffusion for reactants and products, and gas phase heat transfer. The vapor/liquid equilibrium (VLE) and solution speciation was represented in

AspenONE® with the NRTL electrolyte model regressed on the data of Cullinane (2005) by Hilliard (2005). Base case performance of MEA is calculated with the AspenONE® and RateSep™ model by Freguia (2002).

Dow simulated the MDEA/PZ double matrix case using an in-house simulation package, ProComp (v.8.0.6.0).

Trimeric simulated the compression system and ancillary systems (e.g., steam desuperheating, cooling water) using Design II WinSim (v9.33), a commercial process simulator. The Peng-Robinson equation of state was the thermodynamic model used for the inlet gas blower and direct contact cooler as well as the compression unit operations; ASME steam tables were used for the steam system simulations.

Using stream and unit operations reports from the various simulators, Trimeric prepared overall heat and material balances for the three cases. Additional details on the process simulations are provided in Section 3.

### 2.3.3 Equipment Sizing

After completing the heat and material balances, Trimeric prepared equipment specifications, sized and selected equipment. Sections 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 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 80% capacity factor was used for the economic analyses.
- The $CO_2$ capture system installation is 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 15.2 MPa (2200 psia) for transport and injection at an off-site location.
- Dehydration is included for all cases.

Economic metrics, such as the cost per tonne $CO_2$ avoided and the effect of $CO_2$ removal systems on the cost of electricity, were developed and are presented in Section 6.

**References (Section 2)**


Department of Energy (DOE) National Energy Technology Laboratory (NETL). "Carbon Capture and Sequestration Systems Analysis Guidelines", April 2005.

Rochelle, G.T., G.S. Goff, J.T. Cullinane, and S. Freguia, "Research Results for $CO_2$ Capture from Flue Gas by Aqueous Absorption/Stripping," Proceedings of the Laurance Reid Gas Conditioning Conference, February 25-27, 2002.

Freguia, S., "Modeling of $CO_2$ Removal from Flue Gases with Monoethanolamine," M.S. Thesis, The University of Texas at Austin, 2002.

Hilliard, M., "Thermodynamics of Aqueous Piperazine/Potassium Carbonate/Carbon Dioxide Characterized by the Electrolyte NRTL Model within Aspen Plus®," M.S. Thesis, Department of Chemical Engineering, The University of Texas at Austin (2005).

Oyenekan, B. "Modeling of Strippers for CO2 Capture by Aqueous Amines," Ph.D. Dissertation, The University of Texas at Austin, 2007.
*Note: Table 3-5 in the dissertation shows estimated energy requirements for solvent-configuration systems used in the screening study of this project.*

Rochelle, G.T., et al. "$CO_2$ Capture by Absorption with Potassium Carbonate." Second Quarterly Report 2007. DOE Award # DE-FC26-02NT41440,

# 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 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

Process simulations were divided into four "blocks":

- Inlet gas train (inlet gas blower, inlet direct contact cooler)
- $CO_2$ capture train
- $CO_2$ compression train
- Steam system

Trimeric used WinSim's Design II, version 9.33, to simulate the inlet gas train, the $CO_2$ compression train, and the steam system for all cases. UT developed the primary process simulations for MEA- and KPIP-based $CO_2$ capture trains using Aspen Technology Inc.'s AspenOne® 2006 with the RateSep™ module for modeling the absorber and the stripper. Dow used an in-house process simulator package, ProComp, version 8.0.6.0, for the MDEA / PZ double matrix case. 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 of the inlet gas train was a flue gas stream exiting a wet FGD scrubber. The output of the inlet gas train simulation was the input for the primary $CO_2$ capture simulations conducted by UT and Dow. These simulations included the entire amine system, which consists of an absorber, regenerator, associated process heat exchangers and pumps. The outputs from the UT and Dow simulations were used as inputs for Trimeric's $CO_2$ compression train simulations, which included all interstage coolers and separators. $CO_2$ dehydration equipment was not simulated but was included in the capital costs, as described in Section 4. Operating costs for the dehydration unit were estimated to be negligible (\$0.01/MCF $CO_2$ or \$0.19/tonne $CO_2$) in comparison with the overall cost of $CO_2$ avoided (\$67.20 /tonne $CO_2$ for the current base case) (Tannehill, 1994). The simulation terminated with a $CO_2$ product delivered to the battery limits at 15.2 MPa (2200 psia) and approximately 40°C (104˚F).

### 3.1.2 Thermodynamic and Physical Properties Specifications

The details of the MEA and MEA / PZ models developed by UT are described first, followed by a description of the MDEA / PZ model.

The absorber is modeled with RadFrac™ using a RateSep™ model, which is a rate-based model framework in AspenONE®. The stripper is a reboiled column with two equilibrium stages, one of which is a reboiler The model uses instantaneous reactions in the stripper due to the high temperatures present; however, finite reaction rates are required to accurately model the absorber due to the lower temperatures found in that unit operation. The model includes the effects of liquid-phase and gas-phase diffusion resistances for both the absorber and the stripper.

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:

$$H_2O + MEA^+ \longleftrightarrow H_3O^+ + MEA \quad (1)$$

$$2\ H_2O \longleftrightarrow H_3O^+ + OH^- \quad (2)$$

| | | | |
|---|---|---|---|
| $H_2O + HCO_3^-$ | $\longleftrightarrow$ | $H_3O^+ + CO_3^{2-}$ | (3) |
| $H_2O + CO_2 + MEA$ | $\longrightarrow$ | $MEA^+ + HCO_3^-$ | (4) |
| $HCO_3^- + MEA^+$ | $\longrightarrow$ | $H_2O + CO_2 + MEA$ | (5) |
| $H_2O + CO_2 + MEA$ | $\longrightarrow$ | $MEACO_2^- + H_3O^+$ | (6) |
| $MEACOO^- + H_3O^+$ | $\longrightarrow$ | $H_2O + CO_2 + MEA$ | (7) |

Equations one through three are equilibrium equations; equations four through seven are kinetic equations. Equations four and five are amine-catalyzed bicarbonate formation. The rate coefficients are assumed equal to that of the MDEA catalyzed reaction; these coefficients are based on a model provided by Little at al. (1971). For equations six and seven, the rate expression began with the model of Hikita et al. (1977) and was modified according to experimental data provided by Aboudheir (2002). For the 7 m MEA / 2 m PZ double matrix case, 11 m MEA was used to simulate 7 m MEA / 2 m PZ. The rate constant for carbamate formation was increased by a factor of four to represent the rate enhancement provided by 2 m piperazine. The reactions included in the stripper 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 Design Institute for Physical Properties (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 - Henry's Law components included $CO_2$, $N_2$, and $O_2$.

Dow used an in-house process simulation package, ProComp v.8.0.6.0. The Dow simulation for the MDEA / PZ case uses the Electrolyte NRTL model to calculate vapor-liquid equilibrium. The model is regressed using data that Dow has obtained through years of laboratory and field data and that is validated through commercial-scale production and use of their proprietary solvent formulations. Historically, the acid gas treating systems have used low pressure strippers. The use of higher pressure strippers does represent a departure from Dow's typical applications and is an area where some extrapolation from historical VLE data sets is required. The absorber and stripper models account for the effects of mass transfer as well as reaction kinetics. The heat transfer and mass transfer calculations are extensively supported by commercial-scale operations. Additional details of the Dow models are proprietary and cannot be disclosed here.

### 3.1.3 Key Process Simulation Specifications

Process simulation inputs are presented in Tables 3-1 (Metric units) and Table 3-2 (English units). These inputs supplement the design basis presented in Table 2-1.

**Table 3-1. Summary of Process Simulation Inputs (Metric Units)**

| Description | Units | Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| **Equipment Data - Inlet Gas Conditioning Train (Common to all configurations)** | | | | |
| *Inlet Booster Fan* | | | | |
| Flow rate | std m3/s | 558 | = | = |
| Pressure increase | kPa | 10 | = | = |
| Efficiency | % | 75 | = | = |
| *Direct Contact Cooler* | | 0 | = | = |
| Outlet gas temperature | C | 40 | = | = |
| **Equipment Data - CO2 Capture** | | | | |
| *Absorber* | | | | |
| $CO_2$ removal | % | 90 | = | = |
| Approach to flooding | % | 80 | = | = |
| Absorber maximum diameter | m | 12 | = | = |
| Packing type | - | CMR#2 | = | Flexipac |
| Height of packing | m | 23 | = | 15.2 |
| *Rich Amine Pump* | | | | |
| Pressure increase | kPa | 483 | = | = |
| Efficiency | % | 65 | = | = |
| *Rich Amine Carbon Filter* | | | | |
| Slipstream fraction of rich circulation rate | % | 15 | = | = |
| Total filtration allowable pressure drop | kPa | 69 | = | = |
| *Particulate Filter* | | | | |
| Slipstream fraction of rich circulation rate | % | 15 | = | = |
| *Rich Amine High Pressure Booster Pump* | | | | |
| Pressure increase | kPa | - | 107 | 124 |
| Efficiency | % | - | 65 | 65 |
| *Rich/Lean Amine Exchanger* | | | | |
| Cold-side temperature approach | C | 5 | = | = |
| Allowable pressure drop - lean | kPa | 138 | = | = |
| Allowable pressure drop - rich | kPa | 138 | = | = |
| *Rich/Semi-Lean Amine Exchanger* | | | | |
| Cold-side temperature approach | C | - | 5 | 5 |
| Allowable pressure drop - lean | kPa | - | 138 | 138 |
| Allowable pressure drop - rich | kPa | - | 138 | 138 |
| *Low Pressure Stripper* | | | | |
| Bottom Pressure | kPa | 172 | = | = |
| Approach to flooding | % | 80 | = | = |
| Packing type | - | CMR#2 | = | Flexipac 1Y |
| Total height of packing | m | 2 | 5.3 | 13.7 |
| *Low Pressure Reboiler* | | | | |
| Number | - | One per stripper | = | = |
| *Low Pressure Condenser* | | | | |
| Number | - | One per stripper | = | = |
| Process-side outlet temperature | C | 40 | = | = |
| Allowable pressure drop - process | kPa | 14 | = | = |
| Allowable pressure drop - cooling water | kPa | 207 | = | = |
| *Low Pressure Condenser Accumulator* | | | | |
| Number | - | One per condenser | = | = |
| *Low Pressure Stripper Condensate Pump* | | | | |
| Pressure increase | kPa | 207 | 276 | = |
| Efficiency | % | 65 | = | = |

*Note: "=" indicates a value equal to the base case.*

**Table 3-1. Summary of Process Simulation Inputs (Metric Units, continued)**

| Description | Units | Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| *Low Pressure Lean Amine Pump* | | | | |
| Pressure increase | kPa | 414 | = | 345 |
| Efficiency | % | 65 | = | = |
| *Low Pressure Semi-Lean Pump* | | | | |
| Pressure increase | kPa | - | 324 | 296 |
| Efficiency | % | - | 65 | 65 |
| *High Pressure Stripper* | | | | |
| Bottom Pressure | kPa | - | 279 | 296 |
| Approach to flooding | % | - | 80 | 80 |
| Packing type | - | - | CMR#2 | Flexipac 1Y |
| Height of packing | m | - | 1.5 | 12.2 |
| *High Pressure Reboiler* | | | | |
| Number | - | - | One per stripper | One per stripper |
| *High Pressure Stripper Condensate Pump* | | | | |
| Pressure increase | kPa | - | 138 | 69 |
| Efficiency | % | - | 65 | 65 |
| *High Pressure Lean Amine Pump - ELIMINATED* | | | | |
| *Lean Cooler* | | | | |
| Process outlet temperature | C | 40 | = | = |
| Allowable process-side pressure drop | kPa | 69 | = | = |
| Allowable shell-side pressure drop | kPa | 207 | = | = |
| *Semi-lean Cooler* | | | | |
| Process outlet temperature | C | - | 40 | 40 |
| Allowable process-side pressure drop | kPa | - | 69 | 69 |
| Allowable shell-side pressure drop | kPa | - | 207 | 207 |
| **Equipment Data - CO2 Compression** | | | | |
| *Compressors* | | | | |
| Number of stages | - | 4 | 5 | 5 |
| Compressor discharge pressure | kPa | 9653 | = | = |
| Polytropic efficiency | % | 80 | = | = |
| Maximum discharge temperature | C | 149 | = | = |
| *Compressor Pump (last stage)* | | | | |
| Discharge pressure | kPa | 15200 | = | = |
| Efficiency | % | 60 | = | = |
| *Compressor Interstage Coolers* | | | | |
| Type | | Water-cooled shell and tube | = | = |
| Process-side outlet temperature | C | 40 | = | = |
| Allowable process-side pressure drop | kPa | 69 | = | = |
| Allowable shell-side pressure drop | kPa | 207 | = | = |
| *Steam Turbine - $CO_2$ Compressor Driver* | | | | |
| Isentropic efficiency | % | 72 | = | = |
| Inlet temperature | C | 316 | = | = |
| Inlet pressure | kPa | 1103 | = | = |
| Turbine discharge pressure | kPa | 239 | = | 308 |

*Note: "=" indicates a value equal to the base case.*

# Table 3-2. Summary of Process Simulation Inputs (English Units)

| Description | Units | Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| **Equipment Data - Inlet Gas Conditioning Train (Common to all configurations)** | | | | |
| *Inlet Booster Fan* | | | | |
| Flow rate | MMSCFD | 1702 | = | = |
| Pressure increase | psi | 2 | = | = |
| Efficiency | % | 75 | = | = |
| *Direct Contact Cooler* | | 0 | = | = |
| Outlet gas temperature | F | 104 | = | = |
| **Equipment Data - CO2 Capture** | | | | |
| *Absorber* | | | | |
| $CO_2$ removal | % | 90 | = | = |
| Approach to flooding | % | 80 | = | = |
| Absorber maximum diameter | ft | 40 | = | = |
| Packing type | - | CMR#2 | = | Flexipac 1Y |
| Height of packing | ft | 74 | = | 50 |
| *Rich Amine Pump* | | | | |
| Pressure increase | psi | 70 | = | = |
| Efficiency | % | 65 | = | = |
| *Rich Amine Carbon Filter* | | | | |
| Slipstream fraction of rich circulation rate | % | 15 | = | = |
| Total filtration allowable pressure drop | psi | 10 | = | = |
| *Particulate Filter* | | | | |
| Slipstream fraction of rich circulation rate | % | 15 | = | = |
| *Rich Amine High Pressure Booster Pump* | | | | |
| Pressure increase | psi | - | 15.5 | 17.9 |
| Efficiency | % | - | 65 | 65 |
| *Rich/Lean Amine Exchanger* | | | | |
| Cold-side temperature approach | F | 9 | = | = |
| Allowable pressure drop - lean | psi | 20 | = | = |
| Allowable pressure drop - rich | psi | 20 | = | = |
| *Rich/Semi-Lean Amine Exchanger* | | | | |
| Cold-side temperature approach | F | - | 9 | 9 |
| Allowable pressure drop - lean | psi | - | 20 | 20 |
| Allowable pressure drop - rich | psi | - | 20 | 20 |
| *Low Pressure Stripper* | | | | |
| Bottom Pressure | psia | 25 | = | = |
| Approach to flooding | % | 80 | = | = |
| Packing type | - | CMR#2 | = | Flexipac 1Y |
| Total height of packing | ft | 5 | 17.2 | 45 |
| *Low Pressure Reboiler* | | | | |
| Number | - | One per stripper | = | = |
| *Low Pressure Condenser* | | | | |
| Number | - | One per stripper | = | = |
| Process-side outlet temperature | F | 104 | = | = |
| Allowable pressure drop - process | psi | 2 | = | = |
| Allowable pressure drop - cooling water | psi | 30 | = | = |
| *Low Pressure Condenser Accumulator* | | | | |
| Number | - | One per condenser | = | = |
| *Low Pressure Stripper Condensate Pump* | | | | |
| Pressure increase | psi | 30 | 40 | = |
| Efficiency | % | 65 | = | = |

*Note: "=" indicates a value equal to the base case.*

**Table 3-2. Summary of Process Simulation Inputs (English Units, continued)**

| Description | | Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| *Low Pressure Lean Amine Pump* | | | | |
| Pressure increase | psi | 60 | = | 50 |
| Efficiency | % | 65 | = | = |
| *Low Pressure Semi-Lean Pump* | | | | |
| Pressure increase | psi | - | 47 | 43 |
| Efficiency | % | - | 65 | 65 |
| *High Pressure Stripper* | | | | |
| Bottom Pressure | psia | - | 40.5 | 42.9 |
| Approach to flooding | % | - | 80 | 80 |
| Packing type | | CMR#2 | = | Flexipac 1Y |
| Height of packing | ft | - | 4.9 | 40 |
| *High Pressure Reboiler* | | | | |
| Number | - | - | 0 | 0 |
| *High Pressure Stripper Condensate Pump* | | | | |
| Pressure increase | psi | - | 20 | 10 |
| Efficiency | % | - | 65 | 65 |
| *High Pressure Lean Amine Pump - ELIMINATED* | | | | |
| *Lean Cooler* | | | | |
| Process outlet temperature | F | 104 | = | = |
| Allowable process-side pressure drop | psi | 10 | = | = |
| Allowable shell-side pressure drop | psi | 30 | = | = |
| *Semi-lean Cooler* | | | | |
| Process outlet temperature | F | - | 104 | 104 |
| Allowable process-side pressure drop | psi | - | 10 | 10 |
| Allowable shell-side pressure drop | psi | - | 30 | 30 |
| **Equipment Data - CO2 Compression** | | | | |
| *Compressors* | | | | |
| Number of stages | - | 4 | 5 | 5 |
| Compressor discharge pressure | psia | 1400 | = | = |
| Polytropic efficiency | % | 80 | = | = |
| Maximum discharge temperature | F | 300 | = | = |
| *Compressor Pump (last stage)* | | | | |
| Discharge pressure | psia | 2205 | = | = |
| Efficiency | % | 60 | = | = |
| *Compressor Interstage Coolers* | | | | |
| Type | | Water-cooled shell and tube | = | = |
| Process-side outlet temperature | F | 104 | = | = |
| Allowable process-side pressure drop | psi | 10 | = | = |
| Allowable shell-side pressure drop | psi | 30 | = | = |
| *Steam Turbine - $CO_2$ Compressor Driver* | | | | |
| Isentropic efficiency | % | 72 | = | = |
| Inlet temperature | F | 600 | = | = |
| Inlet pressure | psia | 160 | = | = |
| Turbine discharge pressure | psia | 35 | = | 45 |

*Note: "=" indicates a value equal to the base case.*

### 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 two figures present process flow diagrams for the base case and the double matrix $CO_2$ capture trains and associated steam systems. The flow diagrams for the MEA / PZ double matrix and the MDEA / PZ double matrix are identical. The single compressor train has multiple stages, interstage coolers, and separators as indicated by “n” that are not all shown on the diagram for clarity. Similarly, multiple parallel amine absorber and regenerator trains are shown as one train on the diagram. The MEA base case and the MEA / PZ double matrix had four parallel amine trains; the MDEA / PZ double matrix case had eight parallel amine trains.

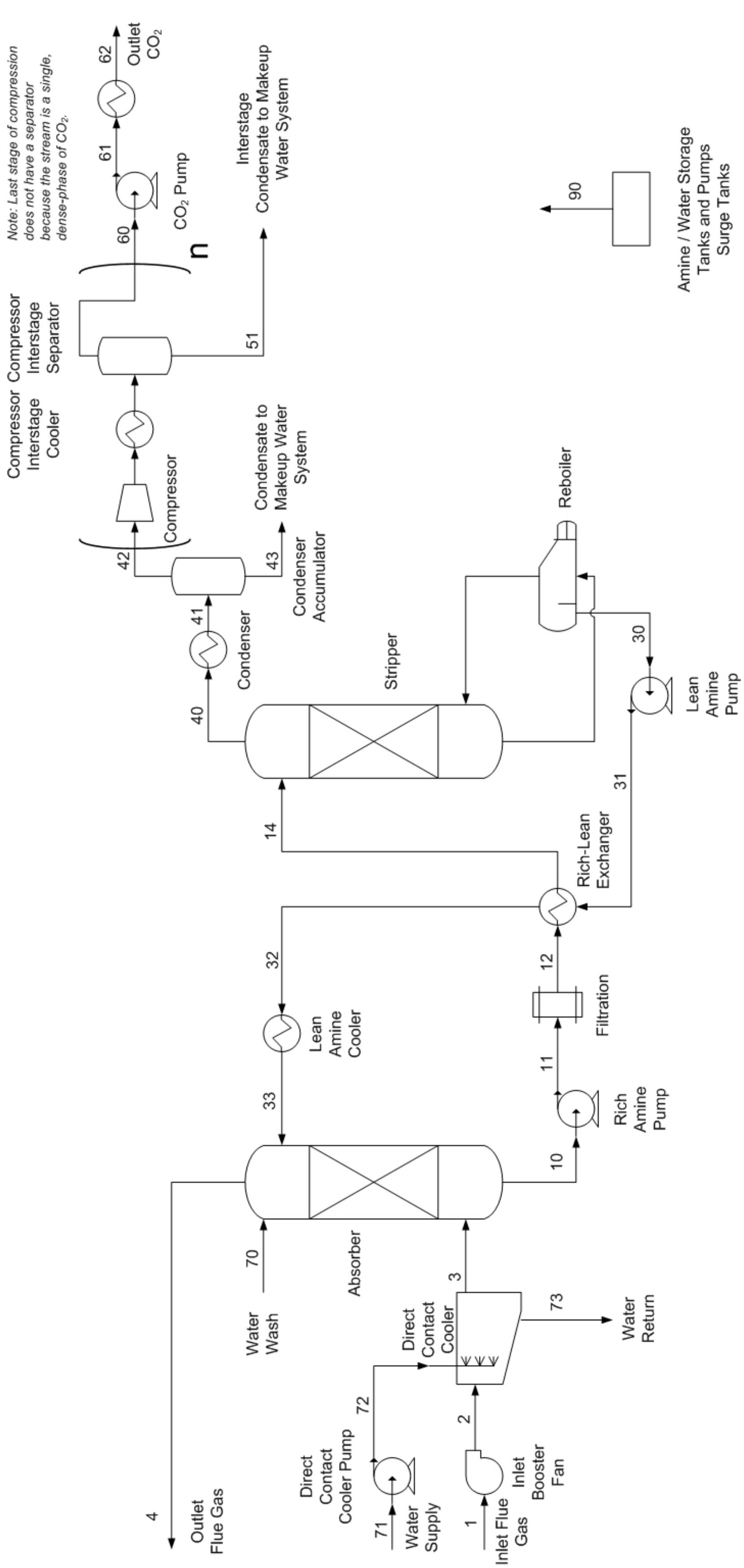


**Figure 3-1. Base Case – Detailed PFD**

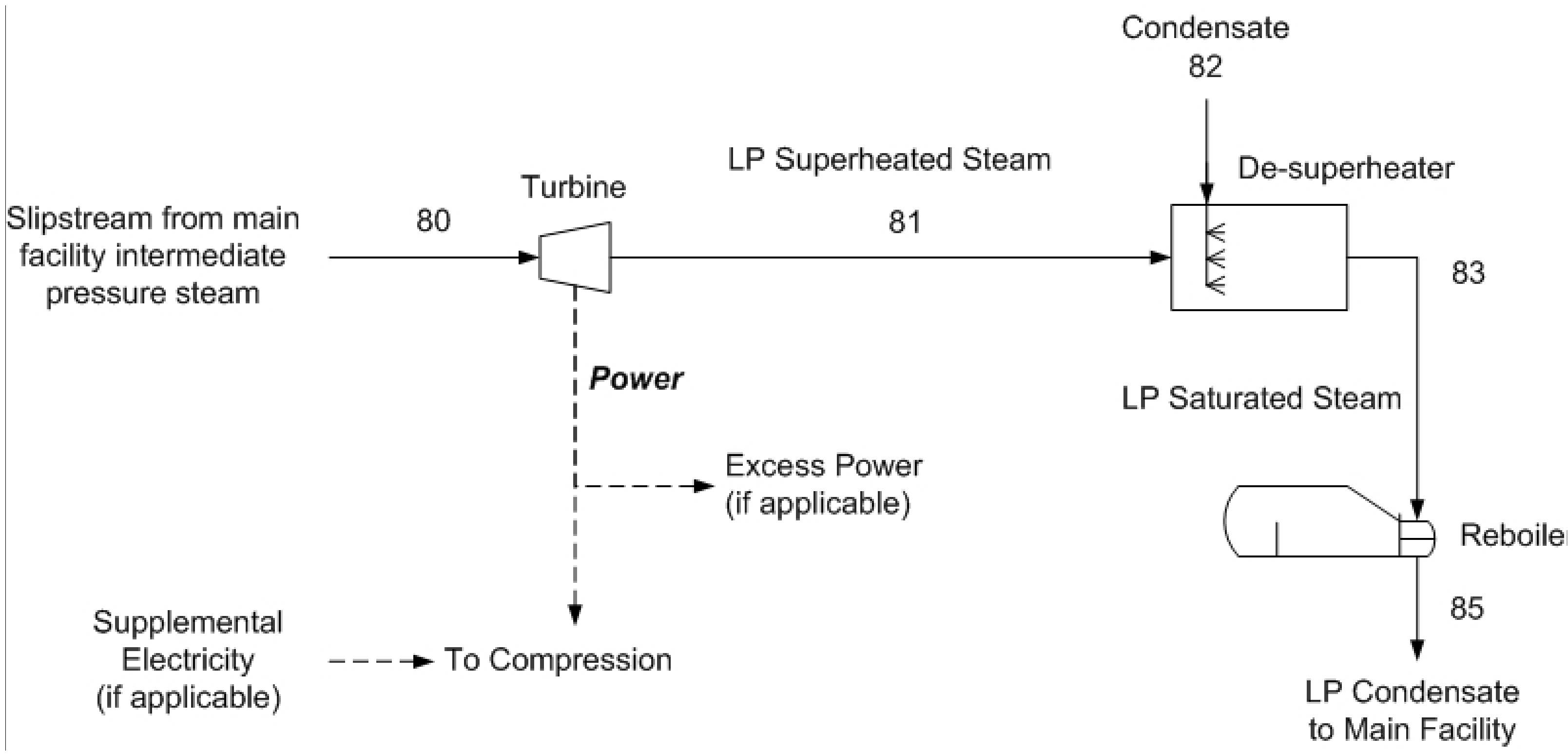


**Figure 3-2. Base Case – Steam System PFD**

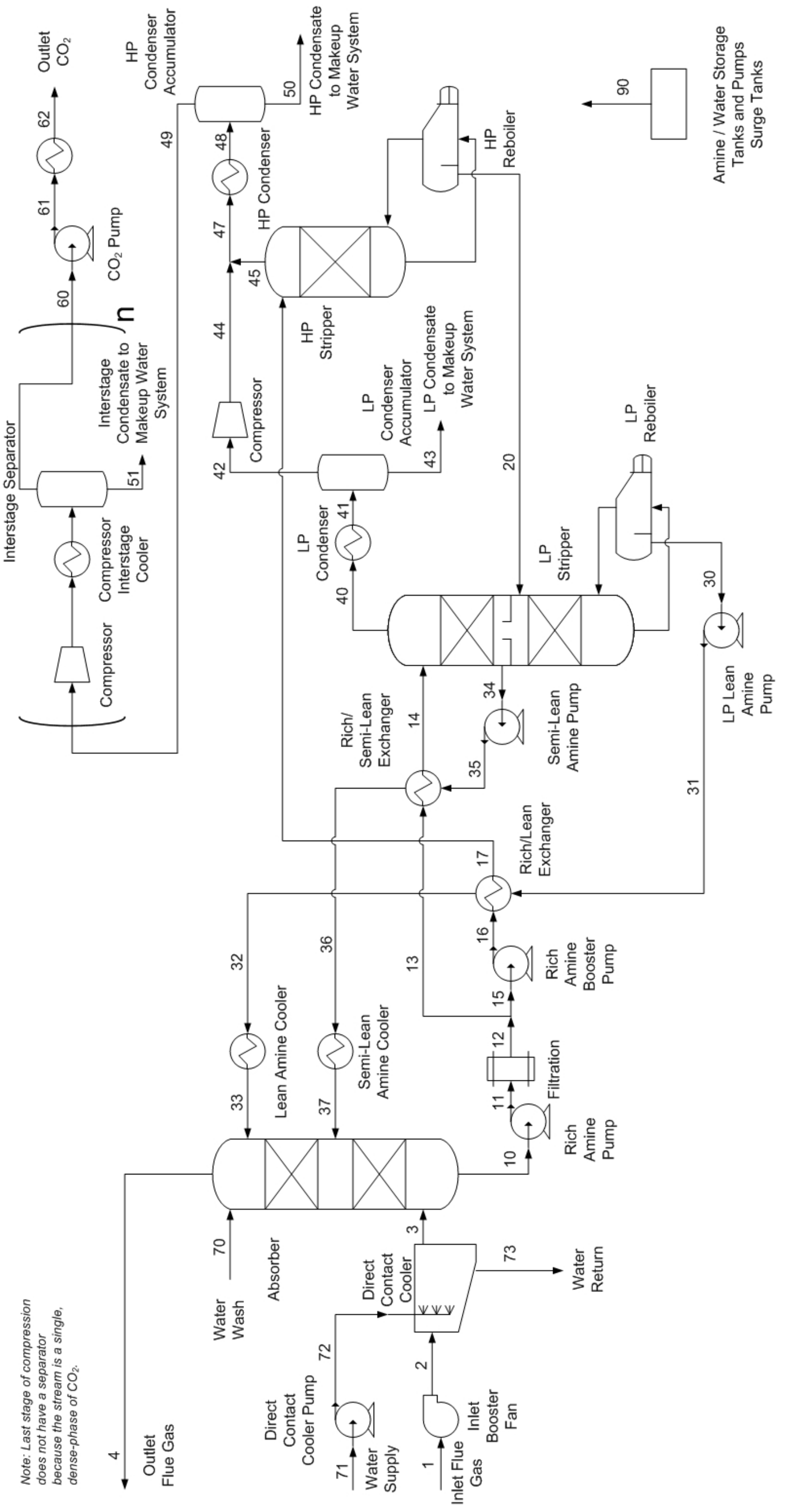


**Figure 3-3. Double Matrix – Detailed PFD**

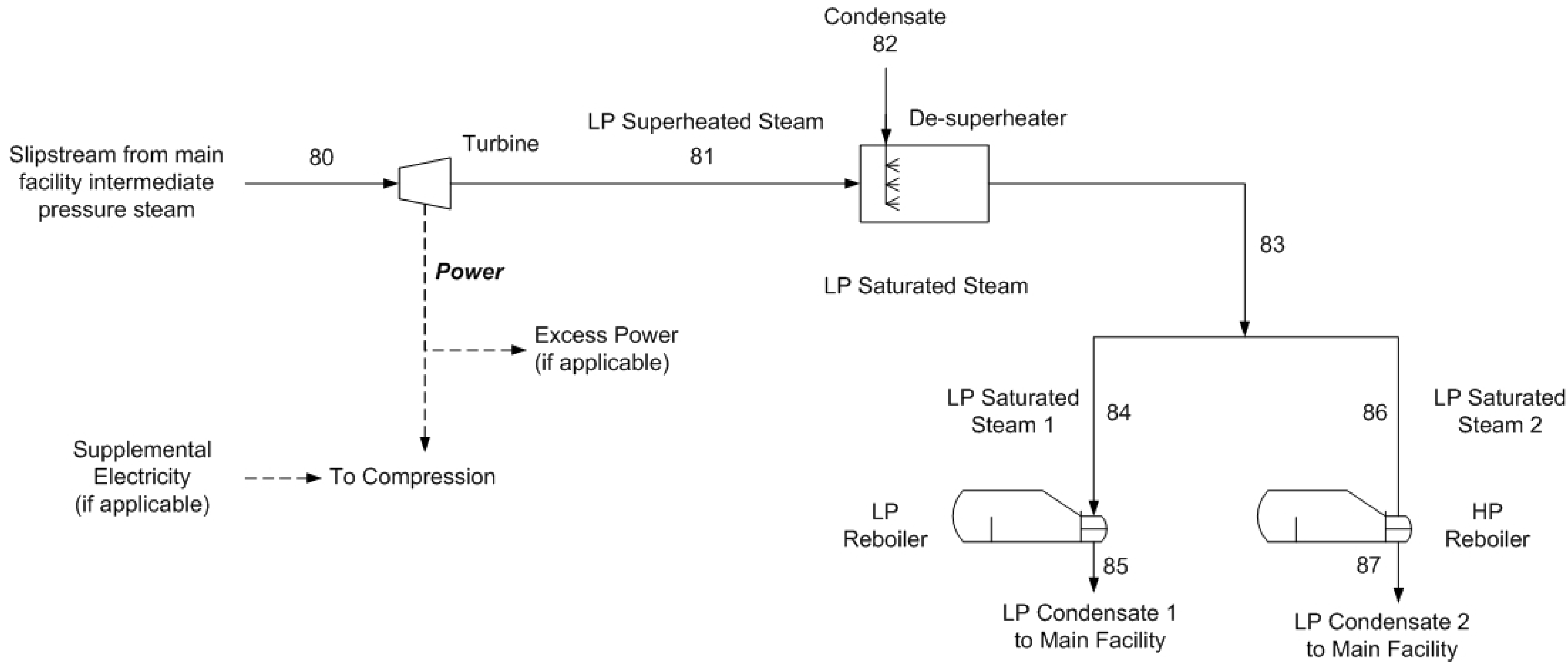


**Figure 3-4. Double Matrix – Steam System PFD**

### 3.2.2 Summary of Process Simulation Results

Key process simulation results are summarized in Table 3-3 (metric units) and 3-4 (English units). For each of the cases, the key simulation parameters (e.g. amine circulation rates, reboiler duties, and compression power) are given. Results for the amine train are given on a per train basis; the MEA cases have four amine trains and the MDEA / PZ case has eight amine trains. The selection of number of trains is determined by maximum absorber size of 40 ft. Several factors contribute to the difference in number and size of absorbers: different packing types, different correlations for flooding, liquid loading (i.e. circulation rates), and different solvent properties. The rich amine pump rate is the overall amine circulation rate. Comparisons between the cases are made in Section 4 in conjunction with the equipment sizing information.

## Table 3-3. Process Simulation Results (Metric)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| Number of inlet gas trains | - | 1 | 1 | 1 |
| Number of CO2 capture trains | - | 4 | 4 | 8 |
| Number of CO2 compression trains | - | 1 | 1 | 1 |
| **Equipment Data - CO2 Capture** | | | | |
| *Absorber* | | | | |
| CO2 removal | % | 90 | 90 | 90 |
| Absorber selected diameter | m | 9.8 | 10.7 | 11.9 |
| Height of packing | m | 22.5 | 22.5 | 15.2 |
| *Rich Amine Pump* | | | | |
| Flow rate per unit | m3/h | 26,129 | 22,931 | 8,456 |
| Brake power per unit | kW/unit | 5,393 | 4,733 | 1,692 |
| *Rich Amine High Pressure Booster Pump* | | | | |
| Flow rate per unit | m3/h per unit | - | 19262 | 6063 |
| Percent of flow to HP stripper | % | - | 84 | 72 |
| Brake power per unit | kW/unit | - | 881 | 311 |
| *Rich/Lean Amine Exchanger* | | | | |
| Duty per unit | kW/unit | 1,197,809 | 715,633 | 351,711 |
| *Rich/Semi-Lean Amine Exchanger* | | | | |
| Duty per unit | kW/unit | - | 109,199 | 143,511 |
| *Low Pressure Stripper* | | | | |
| Bottom Pressure | kPa | 172 | 172 | 172 |
| *Low Pressure Reboiler* | | | | |
| Duty per unit | kW/unit | 485,548 | 193,687 | 35,172 |
| *Low Pressure Condenser* | | | | |
| Duty per unit | kW/unit | 146,712 | 38,969 | 38,707 |
| *Low Pressure Lean Amine Pump* | | | | |
| Flow rate per unit | m3/h | 25,933 | 18,886 | 787 |
| Brake power per unit | kW/unit | 4,588 | 3,341 | 898 |
| *Low Pressure Semi-Lean Pump* | | | | |
| Flow rate per unit | m3/h | - | 3,741 | 342 |
| Flow rate per unit | kW/unit | - | 496 | 470 |
| *High Pressure Stripper* | | | | |
| Bottom Pressure | kPa | - | 279.0 | 296.0 |
| *High Pressure Reboiler* | | | | |
| Duty per unit | kW/unit | - | 192,700 | 234,480 |
| Steam pressure | kPa | - | 240 | 310 |
| *Lean Cooler* | | | | |
| Duty per unit | kW/unit | 352,184 | 288,692 | 82,560 |
| *Semi-lean Cooler* | | | | |
| Duty per unit | kW/unit | - | 55,596 | 37,081 |
| **Equipment Data - CO2 Compression** | | | | |
| *Compressors* | | | | |
| Number of stages | - | 4 | 5 | 5 |
| Total brake power required (total unit) | kW | 40,668 | 38,618 | 35,369 |
| Driver | - | steam | steam | steam and electric |
| Power available from steam | kW | 51,441 | 40,936 | 25,370 |
| Power from electric driver | kW | | 0 | 9,999 |
| Excess available power | kW | 10,773 | 2,317 | 0 |
| *Compressor Pump (last stage)* | | | | |
| Total brake power required (total unit) | kW | 1878 | 1879 | 1883 |
| *Compressor Interstage Coolers* | | | | |
| Total cooler duty | MW(therm)/unit | 75 | 92 | 112 |
| **Equipment Data - Ancillary Equipment** | | | | |
| *Cooling Water System - Utility* | | | | |
| Total | m3/h-unit | 35732 | 29,611 | 16,795 |

## Table 3-4. Process Simulation Results (English)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| Number of inlet gas trains | - | 1 | 1 | 1 |
| Number of CO2 capture trains | - | 4 | 4 | 8 |
| Number of CO2 compression trains | - | 1 | 1 | 1 |
| **Equipment Data - CO2 Capture** | | | | |
| *Absorber* | | | | |
| CO2 removal | | 90 | 90 | 90 |
| Absorber selected diameter | ft | 32.0 | 35 | 39 |
| Height of packing | ft | 74.0 | 74 | 50 |
| *Rich Amine Pump* | | | | |
| Flow rate per unit | gpm | 115,043 | 100,964 | 37,230 |
| Brake power per unit | hp/unit | 7,233 | 6,347 | 2,268 |
| *Rich Amine High Pressure Booster Pump* | | | | |
| Flow rate per unit | gpm per unit | - | 84807 | 26696 |
| Percent of flow to HP stripper | % | - | 84 | 72 |
| Brake power per unit | hp/unit | - | 1,181 | 416 |
| *Rich/Lean Amine Exchanger* | | | | |
| Duty per unit | MMBtu/h-unit | 4,087 | 2,442 | 1,200 |
| *Rich/Semi-Lean Amine Exchanger* | | | | |
| Duty per unit | MMBtu/h-unit | - | 373 | 490 |
| *Low Pressure Stripper* | | | | |
| Bottom Pressure | psia | 25 | 25 | 25 |
| *Low Pressure Reboiler* | | | | |
| Duty per unit | MMBtu/h-unit | 1,657 | 661 | 120 |
| *Low Pressure Condenser* | | | | |
| Duty per unit | MMBtu/h-unit | 501 | 133 | 132 |
| *Low Pressure Lean Amine Pump* | | | | |
| Flow rate per unit | gpm | 114,178 | 83,155 | 3,466 |
| Brake power per unit | hp/unit | 6,153 | 4,481 | 1,205 |
| *Low Pressure Semi-Lean Pump* | | | | |
| Flow rate per unit | gpm | - | 16,473 | 104 |
| Flow rate per unit | hp/unit | - | 665 | 630 |
| *High Pressure Stripper* | | | | |
| Bottom Pressure | psia | - | 40.5 | 42.9 |
| *High Pressure Reboiler* | | | | |
| Duty per unit | MMBtu/h-unit | - | 658 | 800 |
| Steam pressure | psia | - | 35 | 45 |
| *Lean Cooler* | | | | |
| Duty per unit | MMBtu/h-unit | 1,202 | 985 | 282 |
| *Semi-lean Cooler* | | | | |
| Duty per unit | MMBtu/h-unit | - | 190 | 127 |
| **Equipment Data - CO2 Compression** | | | | |
| *Compressors* | | | | |
| Number of stages | | 4 | 5 | 5 |
| Total brake power required (total unit) | hp | 54,536 | 51,788 | 47,431 |
| Driver | | steam | steam | |
| Power available from steam | hp | 68,984 | 54,895 | 34,022 |
| Power from electric driver | hp | | 0 | 13,408 |
| Excess available power | hp | 14,447 | 3,108 | 0 |
| *Compressor Pump (last stage)* | | | | |
| Total brake power required (total unit) | hp | 2518 | 2520 | 2525 |
| *Compressor Interstage Coolers* | | | | |
| Total cooler duty | MMBtu/h-unit | 256 | 315 | 381 |
| **Equipment Data - Ancillary Equipment** | | | | |
| *Cooling Water System - Utility* | | | | |
| Total | gpm per unit | 157325 | 130,373 | 73,948 |

### 3.2.3 Material Balances

Material balances for each of the three cases are given in the Tables 3-5 through 3-7. 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.

Also, the MDEA / PZ solvent formulation is proprietary. To make a mass balance, it was assumed that all amine was MDEA.

**Table 3-5. Material Balance for MEA Base Case**

| Stream Number | - | 1 | 2 | 3 | 4 | 10 | 11 | 12 | 14 | 30 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Inlet Flue Gas | Compressed Gas | Cooled Inlet Gas | Outlet Flue Gas | Rich Amine | Rich Amine - Pump Out | Rich Amine - Filter Out | Rich Amine - Warm to LP Strip | Lean Amine - Hot |
| Temperature | C | 51.3 | 62.8 | 40.0 | 40.2 | 50.8 | 51.0 | 51.1 | 103.9 | 109.0 |
| Pressure | kPa | 101.3 | 111.7 | 111.7 | 102.5 | 111.7 | 594.7 | 525.7 | 387.7 | 172.0 |
| Vapor fraction | - | 1 | 1 | 1 | 1 | | | | | |
| Component molar flow | | | | | | | | | | |
| H2O | kgmol/h | 10,964.4 | 10,964.4 | 5,235.2 | 4,415.4 | 868,341.7 | 868,333.4 | 868,329.6 | 865,242.4 | 858,869.6 |
| CO2 | kgmol/h | 10,493.9 | 10,493.9 | 10,493.6 | 1,049.5 | 6.7 | 6.8 | 6.8 | 662.1 | 226.3 |
| MEA | kgmol/h | | | | 1.1 | 9,843.4 | 9,854.9 | 9,859.8 | 14,689.5 | 27,564.5 |
| N2 | kgmol/h | 59,360.2 | 59,360.2 | 59,360.5 | 59,353.3 | 3.0 | 3.0 | 3.0 | 3.0 | |
| O2 | kgmol/h | 3,941.9 | 3,941.9 | 3,941.1 | 3,943.0 | 0.4 | 0.4 | 0.4 | 0.4 | |
| MEA+ | kgmol/h | | | | | 54,093.3 | 54,090.1 | 54,089.1 | 53,001.9 | 43,951.5 |
| MEACOO- | kgmol/h | | | | | 46,283.2 | 46,274.9 | 46,271.0 | 42,528.5 | 38,695.3 |
| HCO3- | kgmol/h | | | | | 6,677.7 | 6,689.0 | 6,693.8 | 10,213.0 | 5,046.8 |
| CO3-- | kgmol/h | | | | | 565.9 | 562.8 | 561.9 | 129.9 | 104.1 |
| H3O+ | kgmol/h | | | | | | | | | |
| OH- | kgmol/h | | | | | 0.6 | 0.6 | 0.6 | 0.6 | 1.2 |
| HCOO- | kgmol/h | | | | | | | | | |
| Component mass flow | | | | | | | | | | |
| H2O | kg/h | 197,528 | 197,528 | 94,314 | 79,544 | 15,643,425 | 15,643,267 | 15,643,195 | 15,587,582 | 15,472,771 |
| CO2 | kg/h | 461,837 | 461,837 | 461,824 | 46,187 | 295 | 300 | 302 | 29,140 | 9,959 |
| MEA | kg/h | | | | 66 | 601,269 | 601,974 | 602,272 | 897,290 | 1,683,740 |
| N2 | kg/h | 1,662,881 | 1,662,881 | 1,662,890 | 1,662,692 | 83 | 83 | 83 | 83 | |
| O2 | kg/h | 126,137 | 126,137 | 126,112 | 126,172 | 12 | 12 | 12 | 12 | |
| MEA+ | kg/h | | | | | 3,358,714 | 3,358,516 | 3,358,454 | 3,290,947 | 2,728,997 |
| MEACOO- | kg/h | | | | | 4,817,442 | 4,816,574 | 4,816,171 | 4,426,628 | 4,027,643 |
| HCO3- | kg/h | | | | | 407,460 | 408,150 | 408,442 | 623,176 | 307,945 |
| CO3-- | kg/h | | | | | 33,960 | 33,775 | 33,717 | 7,794 | 6,249 |
| H3O+ | kg/h | | | | | | | | | |
| OH- | kg/h | | | | | 10 | 10 | 10 | 10 | 19 |
| HCOO- | kg/h | | | | | | | | | |
| Total molar flow | kgmol/h | 84,761 | 84,761 | 79,031 | 68,762 | 985,816 | 985,816 | 985,816 | 986,471 | 974,459 |
| Total mass flow | kg/h | 2,448,384 | 2,448,384 | 2,345,143 | 1,914,660 | 24,862,665 | 24,862,665 | 24,862,665 | 24,862,665 | 24,237,331 |
| Total volumetric flow | m3/h | 2,253,703 | 2,117,645 | 1,840,649 | 1,746,788 | 26,129 | 26,127 | 26,128 | 26,833 | 25,933 |
| | m3/h std | 2,007,704 | 2,007,704 | 1,871,978 | 1,628,859 | | | | | |
| Molecular weight | kg/kgmol | 28.9 | 28.9 | 29.7 | 27.8 | 25.2 | 25.2 | 25.2 | 25.2 | 24.9 |
| Density | kg/m3 | 1.09 | 1.16 | 1.27 | 1.10 | 951.53 | 951.60 | 951.56 | 926.58 | 934.63 |

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

| Stream Number | - | 31 | 32 | 33 | 40 | 41 | 42 | 43 | 51 | 60 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Lean Amine - Pump Out | Lean Amine - Cool | Lean Amine to Absorber | Stripper Overheads | Stripper Condenser Out | Stripper Condenser Outlet Gas | Stripper Condenser Outlet Liquid | Compression Interstage Condensate | Dense Phase CO2 1 |
| Temperature | C | 109.1 | 56.1 | 40.0 | 101.5 | 40.0 | 40.0 | 40.0 | 40.0 | 40.0 |
| Pressure | kPa | 586.0 | 448.0 | 379.0 | 168.0 | 154.2 | 154.2 | 154.2 | Varies | 9618.2 |
| Vapor fraction | - | | | | 1 | 0.47 | 1 | | | |
| Component molar flow | | | | | | | | | | |
| H2O | kgmol/h | 858,862.6 | 861,100.0 | 861,444.8 | 11,564.2 | 11,555.6 | 483.8 | 11,071.8 | 452.1 | 31.8 |
| CO2 | kgmol/h | 229.0 | 1.7 | 0.3 | 9,461.0 | 9,452.4 | 9,445.8 | 6.9 | 1.6 | 9,444.2 |
| MEA | kgmol/h | 27,577.4 | 24,531.5 | 23,933.1 | 8.6 | | | | | |
| N2 | kgmol/h | | | | 3.0 | 3.0 | 3.0 | | 0.0 | 3.0 |
| O2 | kgmol/h | | | | 0.4 | 0.4 | 0.4 | | 0.0 | 0.4 |
| MEA+ | kgmol/h | 43,948.2 | 44,529.3 | 44,780.9 | | 8.6 | | 8.6 | | |
| MEACOO- | kgmol/h | 38,685.6 | 41,150.5 | 41,497.3 | | | | | | |
| HCO3- | kgmol/h | 5,054.2 | 2,463.2 | 1,867.2 | | 8.6 | | 8.6 | | |
| CO3-- | kgmol/h | 103.6 | 457.1 | 707.4 | | | | | | |
| H3O+ | kgmol/h | | | | | | | | | |
| OH- | kgmol/h | 1.2 | 1.4 | 1.5 | | | | | | |
| HCOO- | kgmol/h | | | | | | | | | |
| Component mass flow | | | | | | | | | | |
| H2O | kg/h | 15,472,656 | 15,512,961 | 15,519,168 | 208,332 | 208,176 | 8,716 | 199,461 | 8,144.7 | 572 |
| CO2 | kg/h | 10,079 | 76 | 14 | 416,379 | 415,999 | 415,708 | 302 | 70.0 | 415,636 |
| MEA | kg/h | 1,684,529 | 1,498,474 | 1,461,921 | 529 | 1 | | 1 | | |
| N2 | kg/h | | | | 83 | 83 | 83 | | 0.0 | 84 |
| O2 | kg/h | | | | 12 | 12 | 12 | | 0.0 | 13 |
| MEA+ | kg/h | 2,728,794 | 2,764,875 | 2,780,496 | | 536 | | 536 | | |
| MEACOO- | kg/h | 4,026,637 | 4,283,192 | 4,319,290 | | 1 | | 1 | | |
| HCO3- | kg/h | 308,398 | 150,301 | 113,933 | | 526 | | 526 | | |
| CO3-- | kg/h | 6,218 | 27,430 | 42,455 | | 0 | | 0 | | |
| H3O+ | kg/h | | | | | | | | | |
| OH- | kg/h | 19 | 25 | 26 | | | | | | |
| HCOO- | kg/h | | | | | | | | | |
| Total molar flow | kgmol/h | 974,462 | 974,235 | 974,233 | 21,037 | 21,029 | 9,933 | 11,096 | 453.7 | 9,479 |
| Total mass flow | kg/h | 24,237,331 | 24,237,331 | 24,237,302 | 625,334 | 625,334 | 424,518 | 200,828 | 8,214.7 | 416,305 |
| Total volumetric flow | m3/h | 25,931 | 25,129 | 24,958 | 386,913 | 166,740 | 166,542 | 227 | | 760 |
| | m3/h std | | | | 498,335 | | 235,295 | | | 224,534 |
| Molecular weight | kg/kgmol | 24.9 | 24.9 | 24.9 | 29.7 | 29.7 | 42.7 | 18.1 | 18.1 | 43.9 |
| Density | kg/m3 | 934.67 | 964.53 | 971.13 | 1.62 | 3.75 | 2.55 | 991.22 | 842.3 | 548.11 |

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

| Stream Number | - | 61 | 62 | 70 | 71 | 72 | 73 | 90 |
|---|---|---|---|---|---|---|---|---|
| Stream Name | - | ESP Outlet | Product | Water Wash (Recycled from Condensate System) | DCC Water Supply | DCC Water Supply - Pump Out | DCC Water Return | Amine Makeup |
| Temperature | C | 58.8 | 40.0 | 40.0 | 29.5 | 29.5 | 40.0 | 40.0 |
| Pressure | kPa | 15237.4 | 15202.9 | 111.7 | 101.3 | 446.1 | 111.7 | 101.3 |
| Vapor fraction | - | | | | | | | |
| Component molar flow | | | | | | | | |
| H2O | kgmol/h | 31.8 | 31.8 | 10,721.5 | 391,446.0 | 391,446.0 | 397,175.0 | |
| CO2 | kgmol/h | 9,444.2 | 9,444.2 | 6.7 | 25.9 | 25.9 | 26.3 | |
| MEA | kgmol/h | | | | | | | 1.4 |
| N2 | kgmol/h | 3.0 | 3.0 | | 3.2 | 3.2 | 3.2 | |
| O2 | kgmol/h | 0.4 | 0.4 | | 54.1 | 54.1 | 54.9 | |
| MEA+ | kgmol/h | | | 8.4 | | | | |
| MEACOO- | kgmol/h | | | | | | | |
| HCO3- | kgmol/h | | | 8.4 | | | | |
| CO3-- | kgmol/h | | | | | | | |
| H3O+ | kgmol/h | | | | | | | |
| OH- | kgmol/h | | | | | | | |
| HCOO- | kgmol/h | | | | | | | |
| Component mass flow | | | | | | | | |
| H2O | kg/h | 572 | 572 | 193,151.4 | 7,052,040 | 7,052,040 | 7,155,247 | |
| CO2 | kg/h | 415,636 | 415,636 | 292.9 | 1,139 | 1,139 | 1,156 | |
| MEA | kg/h | | | 0.7 | | | | 82.7 |
| N2 | kg/h | 84 | 84 | | 89 | 89 | 90 | |
| O2 | kg/h | 13 | 13 | | 1,732 | 1,732 | 1,757 | |
| MEA+ | kg/h | | | 519.2 | | | | |
| MEACOO- | kg/h | | | 1.1 | | | | |
| HCO3- | kg/h | | | 509.4 | | | | |
| CO3-- | kg/h | | | 0.1 | | | | |
| H3O+ | kg/h | | | | | | | |
| OH- | kg/h | | | | | | | |
| HCOO- | kg/h | | | | | | | |
| Total molar flow | kgmol/h | 9,479 | 9,479 | 10,744.9 | 391,530 | 391,530 | 397,260 | 1.4 |
| Total mass flow | kg/h | 416,305 | 416,304 | 194,474.7 | 7,055,010 | 7,055,010 | 7,158,260 | 82.7 |
| Total volumetric flow | m3/h | 710 | | 219.4 | 8,341 | 8,341 | 8,525 | |
| | m3/h std | 224,534 | | | | | | |
| Molecular weight | kg/kgmol | 43.9 | 43.9 | 18.1 | 18.0 | 18.0 | 18.0 | 61.1 |
| Density | kg/m3 | 586.18 | | 991.22 | 845.77 | 845.77 | 839.68 | |

**Table 3-5. Material Balance for MEA Base Case (steam system)**

| Stream Number | - | 80 | 81 | 82 | 83 | 85 |
|---|---|---|---|---|---|---|
| Stream Name | - | Intermediate Pressure Steam | LP Superheated Steam | Condensate | LP Saturated Steam | LP Condensate |
| Temperature | C | 316 | 184 | 40 | 126 | 126 |
| Pressure | kPa | 1103 | 239 | 446 | 239 | 239 |
| Vapor fraction | - | 1 | 1 | 0 | 1 | 0 |
| Total molar flow | kgmol/h | 42,400 | 42,400 | 2,100 | 44,400 | 44,400 |
| Total mass flow | kg/h | 763,000 | 763,000 | 37,000 | 800,000 | 800,000 |
| Total volumetric flow | m3/h | 183,000 | 663,000 | 37 | 599,000 | 853 |
| Molecular weight | kg/kgmol | 18 | 18 | 18 | 18 | 18 |
| Density | kg/m3 | 4.16 | 1.15 | 992 | 1.34 | 938 |

# Table 3-6. Material Balance for MEA / PZ Double Matrix

| Stream Number | | - | 1 | 2 | 3 | 4 | 10 | 11 | 12 | 13 | 14 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | - | Inlet Flue Gas | Compressed Gas | Cooled Inlet Gas | Outlet Flue Gas | Rich Amine | Rich Amine - Pump Out | Rich Amine - Filter Out | Rich Amine to LP | Warm Rich Amine to LP |
| Temperature | | C | 51.3 | 62.8 | 40.0 | 40.8 | 54.8 | 55.0 | 54.9 | 54.9 | 93.1 |
| Pressure | | kPa | 101.3 | 111.7 | 111.7 | 101.7 | 111.7 | 594.7 | 525.7 | 525.7 | 387.0 |
| Vapor fraction | | - | 1 | 1 | 1 | 1 | | | | | |
| Component molar flow | | | | | | | | | | | |
| | H2O | kgmol/h | 10,964.4 | 10,964.4 | 5,235.2 | 4,118.4 | 619,676.2 | 619,663.2 | 619,665.7 | 99,146.5 | 98,755.8 |
| | CO2 | kgmol/h | 10,493.9 | 10,493.9 | 10,493.6 | 1,049.5 | 12.4 | 12.7 | 12.6 | 2.0 | 82.8 |
| | MEA | kgmol/h | | | | 2.0 | 10,055.3 | 10,070.6 | 10,067.8 | 1,610.8 | 2,192.8 |
| | N2 | kgmol/h | 59,360.2 | 59,360.2 | 59,360.5 | 59,354.6 | 1.7 | 1.7 | 1.7 | 0.3 | 0.3 |
| | O2 | kgmol/h | 3,941.9 | 3,941.9 | 3,941.1 | 3,943.2 | 0.2 | 0.2 | 0.2 | 0.0 | 0.0 |
| | MEA+ | kgmol/h | | | | | 60,160.2 | 60,158.0 | 60,158.4 | 9,625.4 | 9,514.9 |
| | MEACOO- | kgmol/h | | | | | 53,053.2 | 53,040.0 | 53,042.4 | 8,486.8 | 8,015.4 |
| | HCO3- | kgmol/h | | | | | 6,526.1 | 6,540.8 | 6,538.1 | 1,046.1 | 1,466.6 |
| | CO3-- | kgmol/h | | | | | 290.4 | 288.5 | 288.9 | 46.2 | 16.4 |
| | H3O+ | kgmol/h | | | | | | | | | |
| | OH- | kgmol/h | | | | | 0.2 | 0.2 | 0.2 | 0.0 | 0.0 |
| | HCOO- | kgmol/h | | | | | | | | | |
| Component mass flow | | | | | | | | | | | |
| | H2O | kg/h | 197,528 | 197,528 | 94,314 | 74,194 | 11,163,639 | 11,163,407 | 11,163,450 | 1,786,152 | 1,779,114 |
| | CO2 | kg/h | 461,837 | 461,837 | 461,824 | 46,187 | 546 | 558 | 556 | 89 | 3,641 |
| | MEA | kg/h | | | | 119 | 614,214 | 615,149 | 614,976 | 98,396 | 133,941 |
| | N2 | kg/h | 1,662,881 | 1,662,881 | 1,662,890 | 1,662,728 | 47 | 47 | 47 | 8 | 8 |
| | O2 | kg/h | 126,137 | 126,137 | 126,112 | 126,177 | 7 | 7 | 7 | 1 | 1 |
| | MEA+ | kg/h | | | | | 3,735,412 | 3,735,279 | 3,735,305 | 597,649 | 590,788 |
| | MEACOO- | kg/h | | | | | 5,522,096 | 5,520,727 | 5,520,976 | 883,357 | 834,289 |
| | HCO3- | kg/h | | | | | 398,205 | 399,104 | 398,938 | 63,830 | 89,486 |
| | CO3-- | kg/h | | | | | 17,424 | 17,313 | 17,335 | 2,773 | 987 |
| | H3O+ | kg/h | | | | | | | | | |
| | OH- | kg/h | | | | | 4 | 4 | 4 | 1 | 1 |
| | HCOO- | kg/h | | | | | | | | | |
| Total molar flow | | kgmol/h | 84,761 | 84,761 | 79,031 | 68,468 | 749,776 | 749,776 | 749,776 | 119,964 | 120,045 |
| Total mass flow | | kg/h | 2,448,384 | 2,448,384 | 2,345,143 | 1,909,405 | 21,451,593 | 21,451,593 | 21,451,593 | 3,432,256 | 3,432,256 |
| Total volumetric flow | | m3/h | 2,253,703 | 2,117,645 | 1,840,649 | 1,756,692 | 22,931 | 22,931 | 22,931 | 3,669 | 3,734 |
| | | m3/h std | 2,007,704 | 2,007,704 | 1,871,978 | 1,621,879 | | | | | |
| Molecular weight | | kg/kgmol | 28.9 | 28.9 | 29.7 | 27.9 | 28.6 | 28.6 | 28.6 | 28.6 | 28.6 |
| Density | | kg/m3 | 1.09 | 1.16 | 1.27 | 1.09 | 935.47 | 935.50 | 935.50 | 935.50 | 919.13 |

**Table 3-6. Material Balance for MEA / PZ Double Matrix (continued)**

| Stream Number | | - | 15 | 16 | 17 | 20 | 30 | 31 | 32 | 33 | 34 |
|---|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | | - | Rich Amine to Booster Pump | Rich Amine - Booster Out | Rich Amine to HP Strip | HP Lean Amine | Hot Lean LP Lean Amine | Lean Amine - Pump Out | Cool Lean Amine | Lean Amine to Absorber | Semi-Lean Amine to Pump |
| Temperature | | C | 54.9 | 55.0 | 101.7 | 107.2 | 107.2 | 107.3 | 60.0 | 40.0 | 97.2 |
| Pressure | | kPa | 525.7 | 632.7 | 494.0 | 279.0 | 172.0 | 586.0 | 448.0 | 379.0 | 168.0 |
| Vapor fraction | | - | | | | | | | | | |
| Component molar flow | | | | | | | | | | | |
| | H2O | kgmol/h | 520,519.1 | 520,516.8 | 518,212.2 | 518,779.0 | 515,558.9 | 515,552.4 | 517,268.3 | 517,689.5 | 101,306.4 |
| | CO2 | kgmol/h | 10.6 | 10.7 | 857.2 | 680.0 | 255.5 | 259.3 | 2.2 | 0.2 | 62.0 |
| | MEA | kgmol/h | 8,456.9 | 8,459.8 | 12,630.2 | 17,427.9 | 24,963.1 | 24,977.5 | 22,591.2 | 22,008.3 | 2,944.0 |
| | N2 | kgmol/h | 1.4 | 1.4 | 1.4 | | | | | | |
| | O2 | kgmol/h | 0.2 | 0.2 | 0.2 | | | | | | |
| | MEA+ | kgmol/h | 50,533.1 | 50,532.7 | 49,513.3 | 46,088.8 | 41,219.6 | 41,215.5 | 41,628.7 | 41,788.3 | 8,973.3 |
| | MEACOO- | kgmol/h | 44,555.6 | 44,553.2 | 41,402.2 | 40,027.3 | 37,351.6 | 37,341.3 | 39,314.4 | 39,737.8 | 7,813.8 |
| | HCO3- | kgmol/h | 5,492.0 | 5,494.7 | 7,972.1 | 5,942.9 | 3,749.7 | 3,756.4 | 1,884.5 | 1,305.6 | 1,129.1 |
| | CO3-- | kgmol/h | 242.6 | 242.3 | 69.4 | 59.2 | 59.0 | 58.6 | 214.6 | 372.1 | 15.1 |
| | H3O+ | kgmol/h | | | | | | | | | |
| | OH- | kgmol/h | 0.2 | 0.2 | 0.2 | 0.3 | 0.5 | 0.5 | 0.6 | 0.7 | 0.0 |
| | HCOO- | kgmol/h | | | | | | | | | |
| Component mass flow | | | | | | | | | | | |
| | H2O | kg/h | 9,377,299 | 9,377,255 | 9,335,739 | 9,345,950 | 9,287,938 | 9,287,821 | 9,318,732 | 9,326,323 | 1,825,063 |
| | CO2 | kg/h | 467 | 469 | 37,725 | 29,927 | 11,244 | 11,412 | 97 | 9 | 2,728 |
| | MEA | kg/h | 516,579 | 516,752 | 771,498 | 1,064,559 | 1,524,839 | 1,525,720 | 1,379,955 | 1,344,347 | 179,833 |
| | N2 | kg/h | 40 | 40 | 40 | | | | | | |
| | O2 | kg/h | 5 | 5 | 5 | | | | | | |
| | MEA+ | kg/h | 3,137,655 | 3,137,630 | 3,074,335 | 2,861,703 | 2,559,371 | 2,559,115 | 2,584,770 | 2,594,680 | 557,162 |
| | MEACOO- | kg/h | 4,637,624 | 4,637,369 | 4,309,389 | 4,166,286 | 3,887,784 | 3,886,713 | 4,092,088 | 4,136,154 | 813,313 |
| | HCO3- | kg/h | 335,107 | 335,274 | 486,439 | 362,622 | 228,796 | 229,210 | 114,988 | 79,666 | 68,897 |
| | CO3-- | kg/h | 14,561 | 14,540 | 4,166 | 3,550 | 3,537 | 3,518 | 12,875 | 22,328 | 908 |
| | H3O+ | kg/h | | | | | | | | | |
| | OH- | kg/h | 4 | 4 | 4 | 5 | 8 | 8 | 11 | 12 | 1 |
| | HCOO- | kg/h | | | | | | | | | |
| Total molar flow | | kgmol/h | 629,812 | 629,812 | 630,658 | 629,005 | 623,158 | 623,162 | 622,905 | 622,903 | 122,244 |
| Total mass flow | | kg/h | 18,019,339 | 18,019,339 | 18,019,339 | 17,834,601 | 17,503,517 | 17,503,517 | 17,503,517 | 17,503,517 | 3,447,907 |
| Total volumetric flow | | m3/h | 19,262 | 19,262 | 19,696 | 19,458 | 18,886 | 18,886 | 18,361 | 18,198 | 3,741 |
| | | m3/h std | | | | | | | | | |
| Molecular weight | | kg/kgmol | 28.6 | 28.6 | 28.6 | 28.4 | 28.1 | 28.1 | 28.1 | 28.1 | 28.2 |
| Density | | kg/m3 | 935.50 | 935.51 | 914.88 | 916.57 | 926.77 | 926.79 | 953.28 | 961.83 | 921.56 |

## Table 3-6. Material Balance for MEA / PZ Double Matrix (continued)

| Stream Number | - | 35 | 36 | 37 | 40 | 41 | 42 | 43 | 44 | 45 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Semi-Lean Amine - Pump Out | Cool Semi-Lean Amine | Semi-Lean to Absorber | LP Stripper Overheads | LP Condenser Out | LP Condenser Outlet Gas | LP Condenser Outlet Liquid | LP Compressor Out | HP Stripper Overheads |
| Temperature | C | 97.3 | 59.9 | 40.0 | 91.7 | 40.0 | 40.0 | 40.0 | 113.5 | 100.7 |
| Pressure | kPa | 478.0 | 340.0 | 271.0 | 160.5 | 126.0 | 126.0 | 126.0 | 275.0 | 275.0 |
| Vapor fraction | - | | | | 1 | 0.69 | 1 | | 1 | 1 |
| Component molar flow | | | | | | | | | | |
| H2O | kgmol/h | 101,305.4 | 101,672.0 | 101,831.1 | 3,201.5 | 3,200.1 | 370 | 2,830.1 | 370 | 1472.6 |
| CO2 | kgmol/h | 62.7 | 1.6 | 0.2 | 5,854.6 | 5,853.3 | 5,851.8 | 1.4 | 5,851.8 | 3,591.5 |
| MEA | kgmol/h | 2,946.6 | 2,431.2 | 2,239.2 | 1.4 | | | | | 1.7 |
| N2 | kgmol/h | | | | 0.3 | 0.3 | 0.3 | | 0.3 | 1.4 |
| O2 | kgmol/h | | | | 0.0 | 0.0 | | | | 0.2 |
| MEA+ | kgmol/h | 8,972.6 | 9,060.2 | 9,091.7 | | 1.4 | | 1.4 | | |
| MEACOO- | kgmol/h | 7,812.1 | 8,239.8 | 8,400.4 | | | | | | |
| HCO3- | kgmol/h | 1,130.3 | 737.2 | 547.9 | | 1.4 | | 1.4 | | |
| CO3-- | kgmol/h | 15.1 | 41.6 | 71.7 | | | | | | |
| H3O+ | kgmol/h | | | | | | | | | |
| OH- | kgmol/h | 0.0 | 0.1 | 0.1 | | | | | | |
| HCOO- | kgmol/h | | | | | | | | | |
| Component mass flow | | | | | | | | | | |
| H2O | kg/h | 1,825,044 | 1,831,650 | 1,834,515 | 57,675 | 57,651 | 6,666 | 50,985 | 6,666 | 26,529 |
| CO2 | kg/h | 2,758 | 69 | 7 | 257,662 | 257,602 | 257,537 | 63 | 257,537 | 158,062 |
| MEA | kg/h | 179,988 | 148,510 | 136,775 | 84 | 0 | | 0.1 | | 103 |
| N2 | kg/h | | | | 8 | 8 | 8 | | 8 | 39 |
| O2 | kg/h | | | | 1 | 1 | | | | 6 |
| MEA+ | kg/h | 557,115 | 562,554 | 564,514 | | 85 | | 85 | | |
| MEACOO- | kg/h | 813,129 | 857,650 | 874,359 | | 0 | | 0 | | |
| HCO3- | kg/h | 68,966 | 44,981 | 33,431 | | 84 | | 84 | | |
| CO3-- | kg/h | 904 | 2,493 | 4,303 | | | | | | |
| H3O+ | kg/h | | | | | | | | | |
| OH- | kg/h | 1 | 1 | 1 | | | | | | |
| HCOO- | kg/h | | | | | | | | | |
| Total molar flow | kgmol/h | 122,245 | 122,184 | 122,182 | 9,058 | 9,056 | 6,222 | 2,834 | 6,222 | 5,067 |
| Total mass flow | kg/h | 3,447,907 | 3,447,907 | 3,447,907 | 315,430 | 315,430 | 264,211 | 51,217 | 264,211 | 184,740 |
| Total volumetric flow | m3/h | 3,741 | 3,671 | 3,644 | 170,059 | 127,880 | 127,752 | 52 | 72,219 | 56,607 |
| | m3/h std | | | | 214,564 | | 147,391 | | 147,391 | 120,038 |
| Molecular weight | kg/kgmol | 28.2 | 28.2 | 28.2 | 34.8 | 34.8 | 42.5 | 18.1 | 42.5 | 36.5 |
| Density | kg/m3 | 921.57 | 939.20 | 946.20 | 1.85 | 2.47 | 2.07 | 991.36 | 3.66 | 3.26 |

## Table 3-6. Material Balance for MEA / PZ Double Matrix (continued)

| Stream Number | - | 47 | 48 | 49 | 50 | 51 | 60 | 61 | 62 | 70 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | LP/HP Vapor Mix | Cool LP/HP Mix | Cool LP/HP Vapor | Cool LP/HP Liquid (1st Stage Condensate) | Interstage Condensate (Stages 2, 3, and 4) | Dense Phase CO2 1 | ESP Pump Out | CO2 Product | Water Wash (Recycled from Condensate System) |
| Temperature | C | 107.8 | 40.0 | 40.0 | 40.0 | 40.0 | 40.0 | 58.9 | 40.0 | 40.0 |
| Pressure | kPa | 275.0 | 261.2 | 261.2 | 261.2 | Varies | 9618.2 | 15237.4 | 15168.5 | 111.7 |
| Vapor fraction | - | 1 | 0.86 | 1 | | | 1 | 1 | | |
| Component molar flow | | | | | | | | | | |
| H2O | kgmol/h | 1842.6 | 1,840.9 | 274.724 | 1,566.2 | 245.0 | 29.7107 | 29.7107 | 29.7107 | 3,547.2 |
| CO2 | kgmol/h | 9,443.3 | 9,441.6 | 9,441.5 | 0.1 | 1.2 | 9,440.3 | 9,440.3 | 9,440.3 | 1.4 |
| MEA | kgmol/h | 1.7 | | | | | | | | |
| N2 | kgmol/h | 1.7 | 1.7 | 1.7 | 0.0 | 0.0 | 1.7 | 1.7 | 1.7 | |
| O2 | kgmol/h | 0.2 | 0.2 | 0.2 | 0.0 | 0.0 | 0.2 | 0.2 | 0.2 | |
| MEA+ | kgmol/h | | 1.7 | | 1.7 | | | | | 2.3 |
| MEACOO- | kgmol/h | | | | | | | | | |
| HCO3- | kgmol/h | | 1.7 | | 1.7 | | | | | 2.3 |
| CO3-- | kgmol/h | | | | | | | | | |
| H3O+ | kgmol/h | | | | | | | | | |
| OH- | kgmol/h | | | | | | | | | |
| HCOO- | kgmol/h | | | | | | | | | |
| Component mass flow | | | | | | | | | | |
| H2O | kg/h | 33,195 | 33,165 | 4,949 | 28,216 | 4,414.0 | 535 | 535 | 535 | 63,901 |
| CO2 | kg/h | 415,599 | 415,525 | 415,521 | 4 | 53.7 | 415,467 | 415,467 | 415,467 | 62 |
| MEA | kg/h | 103 | | | | | | | | 0 |
| N2 | kg/h | 48 | 48 | 48 | 0 | 0.0 | 48 | 48 | 48 | |
| O2 | kg/h | 6 | 6 | 6 | 0 | 0.0 | 6 | 6 | 6 | |
| MEA+ | kg/h | | 104.7 | | 104.7 | | | | | 145 |
| MEACOO- | kg/h | | | | | | | | | 0 |
| HCO3- | kg/h | | 102.7 | | 102.7 | | | | | 142 |
| CO3-- | kg/h | | | | | | | | | 0 |
| H3O+ | kg/h | | | | | | | | | |
| OH- | kg/h | | | | | | | | | |
| HCOO- | kg/h | | | | | | | | | |
| Total molar flow | kgmol/h | 11,289 | 11,288 | 9,718 | 1,570 | 246 | 9,472 | 9,472 | 9,472 | 3,553.2 |
| Total mass flow | kg/h | 448,951 | 448,951 | 420,524 | 28,427 | 4,468 | 416,056 | 416,056 | 416,055 | 64,252 |
| Total volumetric flow | m3/h | 128,875 | 95,629 | 95,629 | | | 760 | 710 | | 64.8 |
| | m3/h std | 267,428 | 267,389 | 230,206 | 37,182 | | 224,374 | 224,374 | 224,374 | |
| Molecular weight | kg/kgmol | 39.8 | 39.8 | 43.3 | 18.1 | 18.1 | 43.9 | 43.9 | 43.9 | 18.1 |
| Density | kg/m3 | 3.48 | 4.40 | 4.40 | | | 547.45 | 585.60 | | 991.7 |

**Table 3-6. Material Balance for MEA / PZ Double Matrix (continued)**

| Stream Number | - | 71 | 72 | 73 | 90 |
|---|---|---|---|---|---|
| **Stream Name** | **-** | **DCC Water Supply** | **DCC Water Supply - Pump Out** | **DCC Water Return** | **Makeup Amine** |
| Temperature | C | 29.5 | 29.5 | 40.0 | 40.0 |
| Pressure | kPa | 101.3 | 446.1 | 101.3 | 101.3 |
| Vapor fraction | - | | | | |
| Component molar flow | | | | | |
| H2O | kgmol/h | 391,446.0 | 391,446.0 | 397,175.0 | |
| CO2 | kgmol/h | 25.9 | 25.9 | 26.3 | |
| MEA | kgmol/h | | | | 2.4 |
| N2 | kgmol/h | 3.2 | 3.2 | 3.2 | |
| O2 | kgmol/h | 54.1 | 54.1 | 54.9 | |
| MEA+ | kgmol/h | | | | |
| MEACOO- | kgmol/h | | | | |
| HCO3- | kgmol/h | | | | |
| CO3-- | kgmol/h | | | | |
| H3O+ | kgmol/h | | | | |
| OH- | kgmol/h | | | | |
| HCOO- | kgmol/h | | | | |
| Component mass flow | | | | | |
| H2O | kg/h | 7,052,040 | 7,052,040 | 7,155,247 | |
| CO2 | kg/h | 1,139 | 1,139 | 1,156 | |
| MEA | kg/h | | | | 148.9 |
| N2 | kg/h | 89 | 89 | 90 | |
| O2 | kg/h | 1,732 | 1,732 | 1,757 | |
| MEA+ | kg/h | | | | |
| MEACOO- | kg/h | | | | |
| HCO3- | kg/h | | | | |
| CO3-- | kg/h | | | | |
| H3O+ | kg/h | | | | |
| OH- | kg/h | | | | |
| HCOO- | kg/h | | | | |
| Total molar flow | kgmol/h | 391,530 | 391,530 | 397,260 | 2.4 |
| Total mass flow | kg/h | 7,055,010 | 7,055,010 | 7,158,260 | 148.9 |
| Total volumetric flow | m3/h | 8,341 | 8,341 | 8,525 | |
| | m3/h std | | | | |
| Molecular weight | kg/kgmol | 18.0 | 18.0 | 18.0 | 61.1 |
| Density | kg/m3 | 845.77 | 845.77 | 839.68 | |

## Table 3-6. Material Balance for MEA / PZ Double Matrix (steam system)

| Stream Number | - | 80 | 81 | 82 | 83 | 84 | 85 | 85 | 86 |
|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Intermediate Pressure Steam | LP Superheated Steam | Condensate | LP Saturated Steam | LP Saturated Steam 1 (To LP Stripper) | LP Condensate 1 | LP Saturated Steam 2 (To HP Stripper) | LP Condensate 2 |
| Temperature | C | 316 | 184 | 40 | 126 | 126 | 126 | 126 | 126 |
| Pressure | kPa | 1103 | 239 | 446 | 239 | 239 | 239 | 239 | 239 |
| Vapor fraction | - | 1 | 1 | 0 | 1 | 1 | 0 | 1 | 0 |
| Total molar flow | kgmol/h | 33,700 | 33,700 | 1,600 | 35,300 | 17,650 | 17,650 | 17,650 | 17,650 |
| Total mass flow | kg/h | 607,000 | 607,000 | 29,000 | 636,000 | 318,000 | 318,000 | 318,000 | 318,000 |
| Total volumetric flow | m3/h | 146,000 | 528,000 | 29 | 476,000 | 238,000 | 339 | 238,000 | 339 |
| Molecular weight | kg/kgmol | 18 | 18 | 18 | 18 | 18 | 18 | 18 | 18 |
| Density | kg/m3 | 4.16 | 1.15 | 992 | 1.34 | 1.34 | 938 | 1.34 | 938 |

## Table 3-7. Material Balance for MDEA / PZ Double Matrix

| Stream Number | - | 1 | 2 | 3 | 4a | 4b | 10 | 11 | 11b | 12 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Inlet Flue Gas | Compressed Gas | Cooled Inlet Gas | Outlet Flue Gas | Outlet Flue Gas (Corrected for Water Wash) | Rich Amine | Rich Amine - Pump Out | Rich Amine - Control Out | Rich Amine - Filter Out |
| Temperature | C | 51.3 | 62.8 | 40.0 | 56.0 | 52.9 | 49.5 | 49.5 | 49.5 | 49.5 |
| Pressure | kPa | 101.3 | 111.7 | 111.7 | 109.9 | 109.9 | 112.2 | 594.9 | 594.9 | 525.9 |
| Vapor fraction | - | 1 | 1 | 1 | 1 | 1 | | | | |
| Total molar flow | kgmol/h | 84,761 | 84,761 | 79,031 | 74,187 | 73,880 | 273,920 | 273,920 | 284,864 | 284,864 |
| Total volumetric flow | m3/h | 2,253,703 | 2,117,645 | 1,840,649 | 1,848,448 | 1,821,056 | 8,456 | 8,456 | 8,662 | 8,662 |
| Total volumetric flow | m3/h std | 2,007,704 | 2,007,704 | 1,871,978 | 1,757,368 | | | | | |

## Table 3-7. Material Balance for MDEA / PZ Double Matrix (continued)

| Stream Number | - | 13 | 14 | 15 | 16 | 17 | 20 | 30 | 31 | 32 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Rich Amine to LP | Warm Rich Amine to LP | Rich Amine to Booster Pump | Rich Amine - Booster Out | Rich Amine to HP Strip | HP Lean Amine | Hot Lean LP Lean Amine | Lean Amine - Pump Out | Cool Lean Amine |
| Temperature | C | 49.5 | 96.7 | 49.5 | 49.6 | 104.8 | 126.7 | 116.2 | 116.3 | 54.6 |
| Pressure | kPa | 525.9 | 388.0 | 525.9 | 649.6 | 511.7 | 296.0 | 172.4 | 517.1 | 379.2 |
| Vapor fraction | - | | | | | | | | | |
| Total molar flow | kgmol/h | 85,458 | 85,458 | 199,406 | 199,406 | 199,406 | 189,960 | 182,231 | 182,231 | 182,231 |
| Total volumetric flow | m3/h | 2,599 | 43,542 | 6,063 | 6,063 | 33,926 | 6,501 | 6,298 | 6,299 | 5,936 |
| | m3/h std | | | | | | | | | |

### Table 3-7. Material Balance for MDEA / PZ Double Matrix (continued)

| Stream Number | - | 33 | 34 | 35 | 36 | 37 | 40 | 41 | 42 | 43 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Lean Amine to Absorber | Semi-Lean Amine to Pump | Semi-Lean Amine - Pump Out | Cool Semi-Lean Amine | Semi-Lean to Absorber | LP Stripper Overheads | LP Condenser Out | LP Condenser Outlet Gas | LP Condenser Outlet Liquid |
| Temperature | C | 40.0 | 110.3 | 110.3 | 54.5 | 40.0 | 96.5 | 40.0 | 40.0 | 40.0 |
| Pressure | kPa | 310.3 | 158.6 | 468.8 | 330.9 | 262.0 | 157.9 | 144.1 | 144.1 | 144.1 |
| Vapor fraction | - | | | | | | 1 | 0.545 | 1 | |
| Total molar flow | kgmol/h | 182,231 | 86,819 | 86,819 | 86,819 | 86,819 | 6,377 | 6,377 | 3,478 | 2,899 |
| Total volumetric flow | m3/h | 5,867 | 2,827 | 2,827 | 2,678 | 2,649 | 123,357 | 62,483 | 62,382 | 53 |
| | m3/h std | | | | | | 151,061 | | 82,381 | |

### Table 3-7. Material Balance for MDEA / PZ Double Matrix (continued)

| Stream Number | - | 44 | 45 | 47 | 48 | 49 | 50 | 51 | 60 | 61 |
|---|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | LP Compressor Out | HP Stripper Overheads | LP/HP Vapor Mix | Cool LP/HP Mix | Cool LP/HP Vapor | Cool LP/HP Liquid (1st Stage Condensate) | Interstage Condensate (Stages 2, 3, and 4) | Dense Phase CO2 1 | ESP Pump Out |
| Temperature | C | 107.0 | 104.8 | 105.4 | 40.0 | 40.0 | 40.0 | 40.0 | 40.0 | 58.9 |
| Pressure | kPa | 295.0 | 295.0 | 295.0 | 281.2 | 281.2 | 281.2 | Varies | 9618.2 | 15237.4 |
| Vapor fraction | - | 1 | 1 | 1 | 0.75 | 1 | | | 1 | 1 |
| Total molar flow | kgmol/h | 3,478 | 9,438 | 12,916 | 12,916 | 9,717 | 3,199 | 226.9 | 9,490 | 9,490 |
| Total volumetric flow | m3/h | 36,966 | 99,269 | 136,243 | 88,733 | 88,731 | 69 | 2 | 762 | 712 |
| | m3/h std | 82,381 | 223,577 | | | 230,183 | | | 224,809 | 224,809 |

**Table 3-7. Material Balance for MDEA / PZ Double Matrix (continued)**

| Stream Number | - | 62 | 70 | 71 | 72 | 73 | 90 |
|---|---|---|---|---|---|---|---|
| Stream Name | - | CO2 Product | Water Wash (Corrected Makeup plus Recycled Condensate) | DCC Water Supply | DCC Water Supply - Pump Out | DCC Water Return | Makeup (Corrected for Water Wash) |
| Temperature | C | 40.0 | 40.0 | 29.5 | 29.5 | 40.0 | 40.0 |
| Pressure | kPa | 15168.5 | 109.9 | 101.3 | 446.1 | 111.7 | 109.9 |
| Vapor fraction | - | | | | | | |
| Total molar flow | kgmol/h | 9,490 | 10,638.5 | 391,530 | 391,530 | 397,260 | 4,307 |
| Total volumetric flow | m3/h | | | 8,341 | 8,341 | 8,525 | |
| | m3/h std | 224,809 | | | | | |

**Table 3-7. Material Balance for MDEA / PZ Double Matrix (steam system)**

| Stream Number | - | 80 | 81 | 82 | 83 | 84 | 85 | 85 | 86 |
|---|---|---|---|---|---|---|---|---|---|
| Stream Name | - | Intermediate Pressure Steam | LP Superheated Steam | Condensate | LP Saturated Steam | LP Saturated Steam 1 (To LP Stripper) | LP Condensate 1 | LP Saturated Steam 2 (To HP Stripper) | LP Condensate 2 |
| Temperature | C | 316 | 203 | 40 | 134 | 134 | 134 | 134 | 134 |
| Pressure | kPa | 1103 | 308 | 446 | 308 | 308 | 308 | 308 | 308 |
| Vapor fraction | - | 1 | 1 | 0 | 1 | 1 | 0 | 1 | 0 |
| Total molar flow | kgmol/h | 23,800 | 23,800 | 1,300 | 25,100 | 12,550 | 12,550 | 12,550 | 12,550 |
| Total mass flow | kg/h | 429,000 | 429,000 | 23,900 | 452,900 | 59,100 | 59,100 | 393,800 | 393,800 |
| Total volumetric flow | m3/h | 103,000 | 301,000 | 24 | 268,000 | 35,000 | 63 | 233,000 | 423 |
| Molecular weight | kg/kgmol | 18 | 18 | 18 | 18 | 18 | 18 | 18 | 18 |
| Density | kg/m3 | 4.16 | 1.43 | 992 | 1.69 | 1.69 | 931 | 1.69 | 931 |

**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. Equipment is sized for a 500 MW unit. The design basis for the unit was described earlier in Section 2; stream and unit operations data were provided in Section 3. A combination of spreadsheet calculations and simulation tools (Aspen Plus, ProComp, Design II, and PDQ$) were used to help size the equipment in the process. The basis of the study is a single, common inlet gas train; multiple parallel amine units; and a single, common $CO_2$ compression train. The inlet gas train consists of the inlet gas booster fan and direct contact cooler. The two MEA cases had four amine trains, and the MDEA case had eight amine trains.

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. 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 Inlet Gas Blower

The inlet gas blower will increase the pressure of the flue gas to overcome pressure drop through the absorber packing. This blower is quite large and may require alloy materials of construction (e.g., Inconel 625, AL-6XN, 2205). The maximum pressure increase is 1.5 psi, and the design flow rate is 558 std m3/s (1,700 MMSCFD) at a nominal suction pressure of 101 kPa (14.7 psia). We assumed a 75% efficiency for this blower, which yielded a power requirement of 8,370 kW/unit (11,200 hp). This is a very unusual application because of the large volume, large pressure increase, and materials requirements. For this application, either axial or centrifugal blowers would be appropriate. Later studies may consider the cost tradeoffs for placing the blower downstream of a direct contact cooler. The tradeoffs would be between a relaxation of materials requirements for the fan, additional water flow or lower water temperature for the direct contact cooler, and an increase in gas temperature to the amine absorber. Absorber intercooling could also reduce the cooling requirements of the direct contact cooler.

## 4.2 Direct Contact Cooler and Water Pump

The direct contact cooler (DCC) sprays water concurrently and horizontally into the FGD outlet flue gas stream. The DCC water cools the flue gas not by evaporation, but by direct contact, as implied by the name. Water condenses from the flue gas when the appropriate water circulation rate is used. Caustic is added to maintain a pH such that $SO_2$ absorbs and $CO_2$ does not absorb. The $SO_2$ absorbs and is neutralized by the caustic. Sulfuric acid may condense in areas not wetted by the caustic solution; therefore, alloy materials (e.g., Inconel 625, AL-6XN, 2205) may be required for the ductwork in this section. The cooler consists of a spray nozzle grid and a larger section of the duct with a sloped liquid collection area, similar to horizontal FGD scrubbers. Because the water sprays in the same direction as the flue gas, the spray transfers momentum to the flue gas and the outlet gas is actually at a higher pressure than the inlet gas. The degree of momentum transfer is site specific, so it was not considered in sizing the inlet gas blower for this study. However, the momentum transfer could reduce pressure increase requirements of the inlet gas blower. The required water circulation rate is 7,050 m3/h (31,100 gpm). A dedicated cooling tower provides evaporative cooling for the recirculating DCC water.

## 4.3 Absorber

The amine-based sorbent contacts the flue gas and absorbs $CO_2$ inside the absorber vessel. The cross sectional area of the absorber is determined from the flue gas flow rate, a target maximum pressure drop of 1.5 psi, and an 80% approach to flooding. A maximum practical diameter of 12.8 meters (42 feet) was chosen; absorber diameters reported in the literature ranged from 7.9 to 12.8 meters (26 to 42 ft) (Rao, 2004).

The absorber is a vertical, packed column with a water wash section at the top to remove vaporized amine from the overhead stream. The packing consists of two sections of equal height. For the MEA cases (improved base case and MEA / PZ double matrix), the height of the packing is approximately 22.5 meters (74 ft) and was optimized in previous work by the University of Texas (Freguia, 2002). For the MDEA / PZ double matrix case, the total height of packing was 50 ft, which was an initial best guess for the MDEA system. 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, 2004).

## 4.4 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, to account for pressure drop through the lines, and to overcome the operating pressure and height requirements in the stripper. The pressure increase provided by rich amine pump is 483 kPa (70 psi). Flow rate per pump varies according to case. A pump efficiency of 65% was used in the study with 50% sparing of equipment. Stainless steel metal components were selected for the pump.

## 4.5 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 15%) 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. It was assumed that carbon steel vessels could 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.

## 4.6 Rich Amine Booster Pump

The rich amine booster pump is present in the double matrix configuration only. The pump provides additional pressure required to transport the rich amine into the high pressure stripper. For the MEA / PZ double matrix case, the pressure increase is 107 kPa (15.5 psi), and for the MDEA / PZ double matrix case, the pressure increase is 124 kPa (17.9 psi). The other specifications are similar to the rich amine pump. Pump efficiency is 65%, process-wetted materials are 316 stainless steel, and sparing is 50%.

## 4.7 Rich/Lean Exchanger

In the rich/lean exchanger, the rich amine is preheated prior to regeneration by heat exchange with the hot lean amine flowing from the regenerator. For the base case flow scheme, the rich amine is flowing to the sole stripper. For the double matrix flow scheme, the rich amine in this exchanger is flowing to the high pressure stripper. In contrast to previous work, the current study used a temperature approach of 5°C (9°F) on the "cold" end in order to reduce reboiler steam requirements. The previous work used an approach of 10°C (18°F) (Fisher, 2005).

Because of this aggressive temperature approach, plate and frame heat exchangers were selected for the heat exchanger type. Since the plates are generally designed to form channels giving highly 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 relatively small volume and at reduced cost per unit of heat transferred. Because the approach temperature has decreased, the total amount of heat transferred has increased. The corresponding capital cost for the exchanger will increase, but the operating costs due to steam requirements and derating will decrease.

The materials contacted by the rich amine are stainless steel; the materials contacted by the lean side are carbon steel. The allowable pressure drop is 20 psi for both sides of the exchanger. 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. A heat transfer coefficient of 651 Btu/h-ft2-F was used for the plate and frame heat exchangers.

## 4.8 Rich/Semi-Lean Exchanger

The rich/semi-lean exchanger applies only to the double matrix flow scheme. In this exchanger, rich amine is preheated prior to entering the top of the LP stripper upper section by semi-lean amine that exits from the bottom of the LP stripper upper section. The specifications for the rich/semi-lean exchanger are similar to those of the rich/lean exchanger. The temperature approach is 5°C (9°F); the materials contacted by the rich amine are stainless steel; the materials contacted by the semi-lean side are carbon steel; and the allowable pressure drop for both sides is 20 psi. Plate and frame heat exchangers were selected for this exchanger as well.

## 4.9 Regeneration

Regeneration of the rich amine solution involves one or more stripper columns and reboiler sections. Each of these areas is discussed below.

### 4.9.1 Stripper

The main function of any amine stripper is to remove $CO_2$ from rich amine solution by steam stripping. The absorption reactions are reversed with heat supplied by a reboiler. In the base case, only one stripper is present. The rich solution flows down through the stripper, which is typically a packed column. Steam rising up through the column strips the $CO_2$ from the amine solution. The base case reboiler pressure is 172 kPa (25.0 psia).

In the double matrix case, both a low pressure (LP) stripper and a high pressure (HP) stripper are present. In the HP stripper, rich amine enters the top of the HP stripper and flows downward through a packed bed. Heat supplied in the HP reboiler generates steam that flows upward and strips out $CO_2$. High pressure lean amine exits from the reboiler and flows to the LP stripper. The LP stripper has two sections: an "upper" section and a "lower" section. The sections may be separate vessels because of their large size. Both sections typically contain packed beds. In the upper section of the LP stripper, rich amine flows downward and contacts vapors that have risen from the lower section. This partially-stripped semi-lean amine is collected at the base of the upper LP stripper section and is pumped back to the middle of the absorber. In the lower section of the LP stripper, HP lean amine enters at the top and flows downward, contacting steam generated by the LP Reboiler. Any vapor that exits the lower section of the LP stripper then flows to the base of the upper section. For the MEA / PZ double matrix case, the LP reboiler operates at the same pressure as the base case reboiler (172 kPa, 25.0 psia), and the HP reboiler operates at 279 kPa (40.5 psia). For the MDEA / PZ double matrix case, the LP Reboiler also operates at 172 kPa (25.0 psia), but the HP reboiler operates at a slightly higher pressure (296 kPa, 42.9 psia) when compared to the MEA / PZ double matrix case.

Each stripper section is sized for 80% approach to flooding in that section; thus, the two sections of the LP stripper may have different diameters. The stripper vessels are carbon steel, and all packing is stainless steel. (GPSA, 1998).

#### 4.9.2 Reboiler

Heat supplied in the reboiler vaporizes part of the lean amine solution and generates steam for stripping. Kettle reboilers are used for this study. Solution flows by gravity from the base of the stripper into the 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. For the base case, one reboiler is present. For the double matrix flow scheme, two reboilers are present: one for the LP stripper and one for the HP stripper. In all cases, the process fluid, lean amine, flows on the shell side of the reboiler, and utility steam from the main power plant flows on the tube side. Utility steam was supplied at 20 psig for the MEA base case and MEA / PZ double matrix case; utility steam was supplied at 30 psig for the MDEA / PZ double matrix case. The log mean temperature differences (LMTD) ranged from 13$^o$C to 22$^o$C (23$^o$F to 40$^o$F). The reboiler tube bundle is stainless steel, and the shell is carbon steel (GPSA, 1998). A heat transfer coefficient of 852 W/m$^2$-K (150 Btu/hr-ft$^2$-°F) was used to size the reboiler tubes when steam is used as the heat source (GPSA, 1998).

### 4.9.3 Stripper Condenser and Accumulator

The stripper condenser cools hot overhead vapors that exit from the top of the stripper. This cooling reduces amine losses and condenses water for subsequent recycling via the absorber water wash stage. Condensed liquids are separated from the $CO_2$ and water vapor in the stripper condenser accumulator, a horizontal vessel located downstream of the condenser. Each stripper condenser has an associated condenser accumulator.

For the base case, one stripper condenser and one condenser accumulator are present. Vapor exiting the condenser accumulator flows to the first stage of compression. Condensed liquids are sent to the makeup water system and, ultimately, are recycled via the water wash system. For the double matrix case, the LP stripper and the HP stripper each have an overhead stripper condenser and condenser accumulator. The HP stripper condenser is also the compressor 1$^{st}$ stage intercooler, and the HP condenser accumulator is also the 1$^{st}$ interstage knockout. Vapor from the LP condenser accumulator flows to the first stage of compression. Hot gas exiting from the first stage compressor combines with HP stripper overheads and flows to the HP condenser, and then on to the HP condenser accumulator. $CO_2$ vapor then exits the HP condenser accumulator and flows to the second stage of compression. Liquids from both accumulators flow to the makeup water system.

The stripper overhead condensers are shell and tube exchangers. Air-cooled exchangers were considered but are not the preferred choice due to the large heat requirements for this application and resulting size of the coolers. Thus, cooling water is the cooling medium. Process materials flow on the tube side, and cooling water flows on the shell side. The tubes are constructed of stainless steel, and the shell is constructed of carbon steel. Cooling water supply temperature is 29°C (85°F), and the return temperature is 43°C (110°F). Process outlet temperature is 40°C (104°F) for all stripper condensers. A heat transfer coefficient of 454 W/m$^2$-K (80 Btu/hr-ft$^2$-°F) was used for the condensers (GPSA, 1998).

The condenser accumulators were sized using the Design II process simulator and assuming a horizontal vessels with minimum liquid residence time of five minutes. The accumulators were constructed from stainless steel.

### 4.9.4 Condensate Pumps

Condensate pumps transport condensed water and recovered amine from the condenser accumulators through the makeup water system and to the absorber water wash. These pumps supply enough power to overcome line losses and the elevation of the absorber water wash. For the base case, the water condensed in the LP condenser is sufficient to meet the absorber water wash requirements. No other condensate or makeup water pumps are required. The condensate pump pressure increase is 30 psi. For the double matrix case, LP condensate pumps accept water from the LP condenser accumulator and increase the pressure by 40 psi. HP condensate pumps accept water from the HP condenser accumulator ($1^{st}$ stage knockout) and increase the pressure by 20 psi. All condensate pumps are centrifugal, are constructed from stainless steel, and have an efficiency of 65%.

## 4.10 Lean Amine Pump

Lean amine solution from the bottom of the stripper is pumped to an elevated pressure to overcome line losses, pressure drops in the rich/lean amine exchanger and lean amine cooler, and the elevation at the top of the absorber. For the base case, the lean amine pump accepts amine from the sole reboiler and the pressure increase is 414 kPa (60 psi). For the double matrix cases, the pump accepts liquid from the LP stripper, and the pressure increases are 414 kPa (60 psi) for the MEA / PZ double matrix and 345 kPa (50 psi) for the MDEA / PZ double matrix. A pump efficiency of 65% was used with 50% sparing of equipment. Pumps are constructed from stainless steel.

### 4.11 Semi-Lean Amine Pump

Semi-lean amine solution flows from the base of the LP stripper upper section, and the semi-lean amine pump provides energy to overcome line losses, pressure drops in the rich/semi-lean amine exchanger and semi-lean amine cooler, and the elevation at the middle of the absorber. Semi-lean amine pumps are only present in the double matrix cases. For the double matrix cases, the pump accepts liquid from the LP stripper, and the pressure increases are 324 kPa (47 psi) for the MEA / PZ double matrix and 345 kPa (50 psi) for the MDEA / PZ double matrix. A pump efficiency of 65% was used with 50% sparing of equipment. Pumps are constructed from stainless steel.

### 4.12 Surge Tank

The surge tank for the lean amine solution was sized based on a 15-minute residence time. Carbon steel was selected for the surge tank.

### 4.13 Lean Amine Cooler

After the rich/lean amine exchanger, the lean amine must be further cooled in a trim cooler before it is pumped back into the absorber column. The trim cooler lowers the lean amine temperature to 40$^o$C (104$^o$F) using cooling water in a counter-current, shell and tube exchanger. Higher temperatures can result in excessive amine evaporative loss and decreased acid gas absorption effectiveness. The exchanger shell is carbon steel, and the tubes are stainless steel. A heat transfer coefficient of 795 W/m$^2$-K (140 Btu/hr-ft$^2$-°F) was used to size the exchanger (GPSA, 1998).

### 4.14 Semi-Lean Amine Cooler

The semi-lean amine cooler is present only in the double matrix flow scheme. As with the lean amine, the semi-lean amine requires trim cooling after cross-exchange in the rich/semi-lean amine exchanger and prior to entering the absorber between the two packed beds. The semi-lean cooler is water cooled with a process outlet temperature of 40°C (104°F). The exchanger shell is carbon steel, and the tubes are stainless steel. A heat transfer coefficient of 795 W/$m^2$-K (140 Btu/hr-$ft^2$-°F) was used to size the exchanger (GPSA, 1998).

### 4.15 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 (1400 psia), at which point the supercritical $CO_2$ forms a dense liquid-like phase. Then, the $CO_2$ is pumped with multistage centrifugal pumps to 15.2 MPa (2210 psia) pipeline pressure. The isentropic efficiency for this type of pump is 60%. Process wetted materials are stainless steel. After passing through an aftercooler, the final $CO_2$ product pressure is 15.2 MPa (2200 psia)

- *Axial versus Centrifugal Compression for First Stage.* The total $CO_2$ capture flow rate for the 500 MW base case is approximately 2,780 $m^3$/min (98,000 acfm). For this size range, either a small axial compressor or a large centrifugal compressor could be used (GPSA 1998). 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 maximum temperature limit of 149°C (300°F). For the MEA base case, there were four compression stages and the final pump stage. For both double matrix cases, there were five compression stages and the final pump stage. The compression in the double matrix cases required five stages because the compression ratio of the first stage is set by the pressure of the HP Stripper. The compression ratio for all of the double matrix compression stages was lower than the ratios for the base case configuration.

### 4.16 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. Because this technology is ultimately a retrofit technology, the second approach was taken in this analysis.

Figures 3-2 and 3-4 show the steam subsystem flow diagram for the base case and double matrix flow schemes, respectively. Superheated steam is taken from the power plant at an intermediate pressure of 1103 kPa and 316˚C (160 psia and 600˚F) to provide the necessary reboiler heat for each of the cases. This steam drives the compressor train with a steam turbine, where the steam pressure drops to 239 kPa (34.7 psia) for both MEA cases and 308 kPa (44.7 psia) for the MDEA case. The turbine exhaust steam is superheated and must be desuperheated with condensate to provide saturated steam to the reboilers. Steam condensate that exits the reboilers then returns to the main facility. The flow rate of steam is somewhat fixed by the heat required in the reboiler(s). In cases where the power delivered from the steam turbine is not enough to drive the compressors, an electric motor will provide the remaining compressor load. In cases where the power supplied by the steam turbine exceeded the energy requirements of the compressor, the excess energy was assumed available to drive a generator.

### 4.17 Interstage Coolers

For all cases, water-cooled exchangers were used for interstage compression cooling. The target outlet $CO_2$ temperature is 40$^o$C (104$^o$F) on the tube side of the exchanger based on the availability of cooling water at 29.4$^o$C (85$^o$F). Cooling water flow to the intercoolers is done in parallel. The exchanger shell is constructed from carbon steel, and the tubes are constructed from stainless steel. The heat transfer coefficients were calculated as a function of pressure using data from the literature (GPSA, 1998); values ranged from 230 W/m$^2$-°C (40 Btu/h-ft$^2$-°F) to 1650 W/m$^2$-°C (290 Btu/h-ft$^2$-°F).

### 4.18 Interstage Separators

Separators are required to separate condensed liquids from the gas downstream of the interstage coolers. The separators were sized with Design II 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 constructed of stainless steel.

### 4.19 Makeup Systems

Because of vaporization losses it is usually necessary to add make-up amine and water to maintain the desired solution strength. The makeup requirement depends on a number of factors such as the reboiler temperature, the stripper condenser temperatures, the compressor interstage cooler temperatures, and the outlet flue gas temperature. 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 caustic in the direct contact cooler is assumed to capture much of the remaining $SO_x$ , this study assumed that some minimal amount of $SO_x$ will slip through. Thus, the nominal loss of all solvents was conservatively estimated at 1.5 kg amine/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 interstage cooling is recycled back to the amine unit and the absorbers are equipped with a water wash at the top. The amine makeup tank was sized to hold one month's worth of chemical and the makeup water about one day. One tank serves all four trains. A makeup amine pump was also included. For the MEA base case and the MEA / PZ double matrix case, recycled interstage condensate eliminated the need for makeup water. However, for these two cases, a water holding tank as well as makeup water pumps were included for operational reasons. The MDEA / PZ double matrix case did require makeup water on a regular basis. The makeup water tank was sized for 12 hour supply for this case.

### 4.20 Cooling Water Systems

Two separate cooling towers were included for the unit. One system, the "DCC" system, provides water to the direct contact cooler. The water that circulates through this system will contain caustic and will absorb species from the gas. The second system, the "utility" system, provides cooling water to all other water-cooled exchangers; this water never directly contacts process material. The DCC system will have different needs with regard to materials of construction, cooling tower chemical addition, etc. Therefore, the two systems were isolated. However, the design and costing for both systems was conducted in a similar manner.

Mechanical draft cooling towers are used with cooling water return and supply temperatures of $43^{o}C$ to $29^{o}C$ ($110^{o}F$ to $85^{o}F$). Wet bulb temperature was set at $7\ ^{o}C$ ($45\ ^{o}F$) per the Systems Analysis Guidelines (DOE 2005). For all case, the DCC cooling water flow rate was 7,050 $m^3/h$ (31,000 gpm). The flow rate to the utility cooling towers vary depending on the case. The utility cooling water flow rates were 35,700 $m^3/h$ (157,000 gpm) for the MEA base case; 29,600 $m^3/h$ (130,000 gpm) for the MEA / PZ double matrix; and 16,800 $m^3/h$ (73,900 gpm) for the MDEA / PZ double matrix.

### 4.21 Dehydration Unit

DOE/NETL systems analysis guidelines stipulate that studies such as this one include the cost of dehydrating the $CO_2$, even though in some cases it may not be necessary to dehydrate the $CO_2$. Since this cost is relatively small and the same for all of the options studied, a detailed effort to size and select this equipment was not necessary. Instead, an allowance for the cost of a standard dehydration unit using triethylene glycol was provided for each of the cases studied.

### 4.22 Reclaimer

As degradation products build in the system, solution reclamation will become necessary. Generally, amines may be thermally reclaimed, though other methods are available. The reclaimer system would be used to remove high boiling point degradation products and sludge. For this study, the cost for a thermal reclaimer system was included. The reclaimer design and cost was assumed to be the same for all cases. In a conventional reclamation system, a small slipstream of the amine solution in circulation (0.5 to 3%) would be routed from the reboiler to a batch distillation reclaimer. MEA solvents may be reclaimed by low pressure steam, and the MEA-laden steam can flow directly back into the LP stripper. MDEA solvents are typically reclaimed under vacuum conditions. Vacuum reclamation is usually conducted on a contract basis. (Kohl and Nielsen, 1997) However, because the degradation rates and solvent losses are not well-defined, this study assumed equivalent solvent reclamation costs for all three cases.

### 4.23 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.

### 4.24 Equipment Comparison for Cases

Tables 4-1 and 4-2 show a comparison of the equipment size requirements for the various cases in this study. Table 4-1 is in metric units, and Table 4-2 is in English units. The tables show the major equipment used in each case along with a brief description of the key sizing parameters. The main differences between the cases are discussed below.

*Significant Equipment Differences:*

- The base case and double matrix configurations have different major equipment lists.

- The MEA cases use a higher circulation rate and do not strip the solvent as lean as the MDEA / PZ case does. The MDEA / PZ solvent has a lower heat of desorption and a lower circulation rate. The MEA base case circulation rate is 26,100 $m^3$/h (115,000 gpm) for the entire unit (all trains); the MEA / PZ double matrix circulation rate is 12% lower at 22,900 $m^3$/h (101,000 gpm), and the MDEA / PZ double matrix circulation rate is 67% lower at 8,500 $m^3$/h (37,000 gpm). The difference in circulation rate affects most unit operations in the $CO_2$ capture trains. The lower circulation rate of the MDEA / PZ double matrix case cannot maintain the flue gas temperature near 40 C (104 F); the outlet flue gas is 53 C (127 F). The hotter outlet gas carries more water, and makeup water is required to supplement condensate recycled from the interstage coolers.

- In the MEA / PZ double matrix, the rich amine booster pumps send 84% of the rich amine to the HP stripper; in the MDEA / PZ double matrix, 70% of the rich amine is pumped to the HP stripper.

  The MEA base case and MEA / PZ double matrix case have four amine trains; the MDEA / PZ double matrix case has eight amine trains. This is primarily due to limits on the absorber diameter. The MEA base case absorber diameter is 9.8 m (32 ft); the MEA / PZ double matrix is 10.7 m (34 ft), and the MDEA / PZ double matrix is 11.9 m (39 ft). The differences in number and diameter of absorbers results from different packing types, flooding correlations, liquid loading, and solvent properties.

- The absorbers in the two MEA cases use 22.5 m (75 ft) of packing, and the MDEA absorbers use 15.2 m (50 ft) of packing. The MEA base case stripper needs minimal packing in the stripper; just enough for one to two equilibrium stages. An alternative to packing might be considered for this configuration in future studies. The MEA / PZ double matrix LP stripper requires less than 6 m (<20 ft) of packing in the LP stripper and minimal packing (1.5 m, 5 ft) in the HP stripper, whereas the MDEA LP stripper uses 13.7 m (45 ft) in the LP stripper and 14.6 m (40 ft) in the HP stripper. The MEA / PZ double matrix HP stripper might use an alternative to packing in order to achieve the one to two equilibrium stages required for that unit operation.

- The stripper diameters vary with case as well. The base case diameter is 7.9 m (26 ft). The MEA / PZ double matrix LP stripper diameter is 4.0 m (13 ft) for the upper section and 6.1 m (20 ft) for the bottom section, and the HP stripper diameter is 6.7 m (22 ft). In part because the MDEA / PZ double matrix case has eight trains yet handles a similar amount of $CO_2$ in the stripping section, the strippers are much smaller than those for the MEA / PZ double matrix. The MDEA stripper diameters are 3.7 m (12 ft) for the entire LP stripper and 4.6 m (15 ft) for the HP stripper. All strippers were sized for an 80% approach to flooding.

- The compression configurations and sizes differ for each case. The base case has one inlet gas stream that flows from the stripper condenser accumulator to the first stage inlet, and the flow rate is 235,000 std m3/h (199 MMSCFD). The MEA / PZ double matrix has two feed streams to the compression train; the gas from the LP stripper condenser accumulator feeds the 1$^{st}$ stage, and HP stripper overheads combine with the 1$^{st}$ stage outlet gas, where the combined stream flows through the 1$^{st}$ stage cooler and on to the 2$^{nd}$ stage of compression. The 1$^{st}$ stage feed gas is 147,000 std m3/h (125 MMSCFD), which is 63% of the base case flow. The combined 2$^{nd}$ stage inlet gas for the MEA / PZ double matrix is 230,000 std m3/h (195 MMSCFD), which is essentially equivalent to the base case. For the double matrix, the 1$^{st}$ stage outlet pressure is tied to the operating pressure of the HP stripper, and there is a tradeoff between extent of stripping in the HP stripper with compression energy as a function of the 1$^{st}$ stage outlet pressure. The MEA / PZ double matrix 1$^{st}$ stage outlet pressure is less than the base case value. Because of this difference, the double matrix case must use five stages of compression instead of four in order to achieve the ESP suction pressure yet remain below the maximum interstage temperature of 149 C (300 F). The MDEA / PZ double matrix LP stripper operates at a slightly higher pressure than the MEA / PZ double matrix; this reduces the 1$^{st}$ stage standard volumetric flow rate ~45% from 128,000 std m3/h (125 MMSCFD) to 62,000 std m3/h (70 MMSCFD). All of this translates into brake power requirements of 40,700 kW (54,500 hp) for the base case, 38,600 kW (51,800 hp) for the MEA / PZ double matrix, and 35,400 kW (47,400 hp) for the MDEA / PZ double matrix.

- The reboiler sizes varied significantly from case to case. The base case has a single reboiler per train, for a total of four large reboilers, each which transfer 121,000 kW (414 MMBtu/h) with a LMTD of 21 C (37 F). The MEA / PZ double matrix has four LP reboilers that transfer 48,400 kW (165 MMBtu/h) each with a LMTD of 21 C (37 F) and four HP reboilers that transfer 48,200 kW (164 MMBtu/h) each with a LMTD of 22 C (40 F). The MDEA / PZ double matrix has eight LP reboilers that transfer 4,400 kW (15 MMBtu/h) with a LMTD of 19 C (34 F) and eight HP reboilers that transfer 29,300 kW (100 MMBtu/h) with a LMTD of 13 C (23 F). The MDEA / PZ double matrix runs the reboilers hotter and uses a higher pressure steam. Also, the heat transfer is evenly divided between the LP and HP reboilers for the MEA / PZ double matrix, but the MDEA / PZ double matrix transfers 87% of the total heat in the HP reboiler.

- In all cases, the amine cross exchangers duties and corresponding sizes were quite large despite the high heat transfer coefficient attained with the plate and frame heat exchangers. The total area of all amine cross exchangers are largest for the MEA base case (68,200 m$^2$ or 735,000 ft$^2$) and decrease in size for the MEA / PZ double matrix (42,900 m$^2$ or 461,000 ft$^2$) and again for the MDEA / PZ double matrix (16,700 m$^2$ or 180,000 ft$^2$).

*Other:*

- The utility cooling water requirements (excluding the direct contact cooler) are 35,700 $m^3$/h (157,000 gpm) for the MEA base case. The MEA / PZ double matrix uses 17% less cooling water, and the MDEA / PZ double matrix cases uses 53% less cooling water than the base case.

- Essentially the same $CO_2$ pump size is required for each case.

## Table 4-1. Equipment Comparison Table (Metric Units)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| Number of inlet gas trains | - | 1 | 1 | 1 |
| Number of CO2 capture trains | - | 4 | 4 | 8 |
| Number of CO2 compression trains | - | 1 | 1 | 1 |
| **Equipment Data - Inlet Gas Conditioning Train (Common to all configurations)** | | | | |
| *Inlet Booster Fan* | | | | |
| Quantity per unit | - | 1 | 1 | 1 |
| Flow rate | std m3/s | 558 | 558 | 558 |
| Pressure increase | kPa | 10.3 | 10.3 | 10.3 |
| Brake power | kW | 8,365 | 8,365 | 8,365 |
| *Direct Contact Cooler* | | | | |
| Water flow rate | m3/h | 7,053 | 7,053 | 7,053 |
| Outlet gas temperature | C | 40 | 40 | 40 |
| *Direct Contact Cooler Water Pump* | | | | |
| Flow rate | m3/h | 7,053 | 7,053 | 7,053 |
| Brake power | kW/unit | 1,039 | 1,039 | 1,039 |
| *Cooling Water System - DCC* | | | | |
| Duty to cool inlet gas | kW | 85,637 | 85,637 | 85,637 |
| Water Rates | m3/h | 7,053 | 7,053 | 7,053 |
| **Equipment Data - CO2 Capture** | | | | |
| *Absorber* | | | | |
| CO2 removal | % | 90 | 90 | 90 |
| Absorber selected diamter | m | 9.8 | 10.7 | 11.9 |
| Height of packing | m | 22.5 | 22.5 | 15.2 |
| *Rich Amine Pump* | | | | |
| Flow rate per train | m3/h | 6,532 | 5,733 | 1,057 |
| Flow rate per unit | m3/h | 26,129 | 22,931 | 8,456 |
| Brake power per train | kW/train | 1,348 | 1,183 | 211 |
| Brake power per unit | kW/unit | 5,393 | 4,733 | 1,692 |
| *Rich Amine Carbon Filter* | | | | |
| Slipstream fraction of rich circulation rate | % | 15 | 15 | 15 |
| Flow rate per train | m3/h-train | 980 | 860 | 159 |
| Flow rate per unit | m3/h-unit | 3,919 | 3,440 | 1,268 |
| *Particulate Filter* | | | | |
| Slipstream fraction of rich circulation rate | % | 15 | 15 | 15 |
| Flowrate per train | m3/h-train | 980 | 860 | 159 |
| Flow rate per unit | m3/h-unit | 3,919 | 3,440 | 1,268 |
| *Rich Amine High Pressure Booster Pump* | | | | |
| Flow rate per train | m3/h per train | - | 4815 | 758 |
| Flow rate per unit | m3/h per unit | - | 19262 | 6063 |
| Brake power per train | kW/train | - | 220 | 39 |
| Brake power per unit | kW/unit | - | 881 | 311 |
| *Rich/Lean Amine Exchanger* | | | | |
| Duty per train | kW/train | 299,452 | 178,908 | 43,964 |
| Duty per unit | kW/unit | 1,197,809 | 715,633 | 351,711 |
| Heat transfer coefficient | W/m2-C | 3,696 | 3696 | 3696 |
| LMTD | C | 5.1 | 5.3 | 7.8 |
| Area per train | m2 | 17,060 | 9,102 | 1,524 |
| Area per unit | m2 | 68,239 | 36,409 | 12,195 |
| *Rich/Semi-Lean Amine Exchanger* | | | | |
| Duty per train | kW/train | - | 27,300 | 17,939 |
| Duty per unit | kW/unit | - | 109,199 | 143,511 |
| Heat transfer coefficient | W/m2-C | - | 3,696 | 3,696 |
| LMTD | C | - | 4.6 | 8.6 |
| Area per train | m2 | - | 1,619 | 567 |
| Area per unit | m2 | - | 6,477 | 4,538 |

# Table 4-1. Equipment Comparison Table (Metric Units, continued)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| *Low Pressure Stripper* | | | | |
| Bottom Pressure | kPa | 172 | 172 | 172 |
| Packing type | - | CMR#2 | CMR#2 | Flexipac 1Y |
| Diameter | m | 7.9 | 4.0 | 3.7 |
| Height of packing | m | 1.5 | 1.5 | 6.1 |
| T-T height | m | 8 | 8 | 7 |
| Diameter | m | - | 6.1 | 3.7 |
| Height of packing | m | - | 3.75 | 7.62 |
| T-T height | m | - | 12 | 20 |
| *Low Pressure Reboiler* | | | | |
| Number | - | One per stripper | One per stripper | One per stripper |
| Duty per train | kW/train | 121,387 | 48,422 | 4,397 |
| Duty per unit | kW/unit | 485,548 | 193,687 | 35,172 |
| Heat transfer coefficient | W/m2-C | 852 | 852 | 852 |
| Steam pressure | kPa | 240 | 240 | 310 |
| LMTD | C | 21 | 21 | 19 |
| Area | m2 | 6,904 | 2,675 | 277 |
| *Low Pressure Condenser* | | | | |
| Number | - | One per stripper | One per stripper | One per stripper |
| Duty per train | kW/train | 36,678 | 9,742 | 4,838 |
| Duty per unit | kW/unit | 146,712 | 38,969 | 38,707 |
| Heat transfer coefficient | W/m2-C | 454 | 454 | 454 |
| LMTD | C | 28 | 25 | 26 |
| Area per train | m2 | 2,896 | 863 | 404 |
| *Low Pressure Condenser Accumulator* | | | | |
| Number | - | ɔne per condense | ɔne per condense | ɔne per condense |
| Diameter | m | 2.7 | 1.2 | 1.8 |
| Length | m | 11.6 | 4.9 | 7.3 |
| *Low Pressure Stripper Condensate Pump* | | | | |
| Flow rate per train | m3/h per train | 57 | 13 | 7 |
| Flow rate per unit | m3/h per unit | 227 | 52 | 53 |
| Brake power per train | kW/train | 5.0 | 1.5 | 0.6 |
| Brake power per unit | kW/unit | 20.0 | 6 | 5 |
| *Low Pressure Lean Amine Pump* | | | | |
| Flow rate per train | m3/h | 6,483 | 4,722 | 6,298 |
| Flow rate per unit | m3/h | 25,933 | 18,886 | 787 |
| Brake power per train | kW/train | 1,147 | 835 | 112 |
| Brake power per unit | kW/unit | 4,588 | 3,341 | 898 |
| *Low Pressure Semi-Lean Pump* | | | | |
| Flow rate per train | m3/h | - | 935 | 353 |
| Flow rate per unit | m3/h | - | 3,741 | 342 |
| Brake power per train | kW/train | - | 124 | 59 |
| Flow rate per unit | kW/unit | - | 496 | 470 |
| *High Pressure Stripper* | | | | |
| Bottom Pressure | kPa | - | 279.0 | 296.0 |
| Packing type | - | - | CMR#2 | Flexipac 1Y |
| Diameter | m | - | 6.7 | 4.6 |
| T-T Length | m | - | 13.4 | 14.6 |
| Height of packing | m | - | 1.5 | 12.2 |
| *High Pressure Reboiler* | | | | |
| Number | - | | One per stripper | One per stripper |
| Duty per train | kW/train | - | 48,175 | 29,310 |
| Duty per unit | kW/unit | - | 192,700 | 234,480 |
| Heat transfer coefficient | W/m2-C | - | 852 | 852 |
| Steam pressure | kPa | - | 240 | 310 |
| LMTD | C | - | 22 | 13 |
| Area | m2 | - | 2,567 | 2,694 |

## Table 4-1. Equipment Comparison Table (Metric Units, continued)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| *High Pressure Stripper Condensate Pump* | | | | |
| Flow rate per unit | m3/h per unit | - | 28 | 69 |
| Brake power per train | kW/train | - | 1.7 | 1.3 |
| *Lean Cooler* | | | | |
| Duty per train | kW/train | 88,046 | 72,173 | 10,320 |
| Duty per unit | kW/unit | 352,184 | 288,692 | 82,560 |
| Heat transfer coefficient | W/m2-C | 795 | 795 | 795 |
| LMTD | C | 12 | 13 | 11 |
| Area | m2 | 9,540 | 6,789 | 1,190 |
| *Semi-lean Cooler* | | | | |
| Duty per train | kW/train | - | 13,899 | 4,635 |
| Duty per unit | kW/unit | - | 55,596 | 37,081 |
| Heat transfer coefficient | W/m2-C | - | 795 | 795 |
| LMTD | C | - | 13 | 11 |
| Area | m2 | - | 1,310 | 536 |
| *Lean Surge Tank* | | | | |
| Tank capacity | m3 | 1,560 | 1,137 | 11,001 |
| *Makeup Amine Tank* | | | | |
| Tank volume | m3 | 446 | 462 | 447 |
| *Makeup Amine Pumps* | | | | |
| Flow rate per unit | m3/h-unit | 0.66 | 0.7 | 0.7 |
| *Water Tank* | | | | |
| Capacity | m3 | 379 | 379 | 930 |
| *Water Pump* | | | | |
| Flow rate per train | m3/h-train | 23 | 23 | 78 |
| **Equipment Data - CO2 Compression** | | | | |
| *Compressors* | | | | |
| Number of stages | - | 4 | 5 | 5 |
| Compressor discharge pressure | kPa | 9,653 | 9,653 | 9,653 |
| Total brake power required (total unit) | kW | 40,668 | 38,618 | 35,369 |
| Inlet gas flow rate - Stage 1 | m3/h | 166,542 | 127,828 | 62,429 |
| | std m3/h | 235,295 | 147,392 | 82,380 |
| Inlet gas pressure - Stage 1 | kPa | 154 | 126 | 144 |
| Inlet gas flow rate - Stage 2 | m3/h | 61,624 | 95,629 | 88,731 |
| | std m3/h | 228,236 | 230,206 | 230,183 |
| Inlet gas pressure - Stage 2 | kPa | 399 | 261 | 281 |
| Driver | - | steam | steam | steam and electric |
| Power available from steam | kW | 51,441 | 40,936 | 25,370 |
| Balance of power required from electric driver | kW | | 0 | 9,999 |
| Excess available power | kW | 10,773 | 2,317 | 0 |
| *Compressor Pump (last stage)* | | | | |
| Discharge pressure | kPa | 15200 | 15237 | 15200 |
| Total brake power required (total unit) | kW | 1878 | 1879 | 1883 |
| *Compressor Interstage Coolers* | | | | |
| Type | | Water-cooled shell and tube | Water-cooled shell and tube | Water-cooled shell and tube |
| Total cooler duty | MW(therm)/unit | 75 | 92 | 112 |
| *Compressor Interstage Separators* | | | | |
| Total capacity | m3 | 51.0 | 88.9 | 88.9 |
| *Steam Turbine - CO2 Compressor Driver* | | | | |
| Isentropic efficiency | % | 72 | 72 | 72 |
| Inlet temperature | C | 315.6 | 316 | 316 |
| Inlet pressure | kPa | 1,103 | 1103 | 1103 |
| Turbine discharge pressure | kPa | 239 | 239 | 308 |

## Table 4-1. Equipment Comparison Table (Metric Units, continued)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| **Equipment Data - Ancillary Equipment** | | | | |
| *Cooling Water System - Utility* | | | | |
| Water Rates | | | | |
| Stage 1 cooler | m3/h | 902 | 1,680 | 2,939 |
| Stage 2 cooler | m3/h | 836 | 720 | 697 |
| Stage 3 cooler | m3/h | 819 | 677 | 661 |
| Stage 4 cooler | m3/h | 1678 | 711 | 700 |
| Stage 5 cooler | m3/h | 421 | 1,529 | 1,514 |
| Stage 6 cooler | m3/h | - | 420 | 421 |
| Lean amine cooler | m3/h per train | 5484 | 4,496 | 643 |
| Semi-lean amine cooler | m3/h per train | - | 866 | 289 |
| LP stripper condenser | m3/h per train | 2285 | 607 | 301 |
| Total | m3/h-unit | 35732 | 29,611 | 16,795 |

**Table 4-2. Equipment Comparison Table (English Units)**

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| Number of inlet gas trains | - | 1 | 1 | 1 |
| Number of CO2 capture trains | - | 4 | 4 | 8 |
| Number of CO2 compression trains | - | 1 | 1 | 1 |
| **Equipment Data - Inlet Gas Conditioning Train (Common to all configurations)** | | | | |
| *Inlet Booster Fan* | | | | |
| Quantity per unit | | | | |
| Flow rate | MMSCFD | 1702 | 1,702 | 1,702 |
| Pressure increase | psi | 1.5 | 1.5 | 1.5 |
| Brake power | hp | 11,218 | 11,218 | 11,218 |
| *Direct Contact Cooler* | | | | |
| Water flow rate | gpm | 31,054 | 31,054 | 31,054 |
| Outlet gas temperature | F | 104 | 104 | 104 |
| *Direct Contact Cooler Water Pump* | | | | |
| Flow rate | gpm | 31,054 | 31,054 | 31,054 |
| Brake power | hp/unit | 1,393 | 1,393 | 1,393 |
| *Cooling Water System - DCC* | | | | |
| Duty to cool inlet gas | MMBtu/h | 292 | 292 | 292 |
| Water Rates | gpm | 31,054 | 31,054 | 31,054 |
| **Equipment Data - CO2 Capture** | | | | |
| *Absorber* | | | | |
| CO2 removal | | 90 | 90 | 90 |
| Absorber selected diamter | ft | 32.0 | 35 | 39 |
| Height of packing | ft | 74.0 | 74 | 50 |
| *Rich Amine Pump* | | | | |
| Flow rate per train | gpm | 28,761 | 25,241 | 4,654 |
| Flow rate per unit | gpm | 115,043 | 100,964 | 37,230 |
| Brake power per train | hp/train | 1,808 | 1,587 | 284 |
| Brake power per unit | hp/unit | 7,233 | 6,347 | 2,268 |
| *Rich Amine Carbon Filter* | | | | |
| Slipstream fraction of rich circulation rate | | | | 15 |
| Flow rate per train | gpm/train | 4,314 | 3,786 | 698 |
| Flow rate per unit | gpm/unit | 17,256 | 15,145 | 5,584 |
| *Particulate Filter* | | | | |
| Slipstream fraction of rich circulation rate | | | | 15 |
| Flowrate per train | gpm/train | 4314 | 3786 | 698 |
| Flow rate per unit | gpm/unit | 17,256 | 15,145 | 5,584 |
| *Rich Amine High Pressure Booster Pump* | | | | |
| Flow rate per train | gpm per train | - | 21202 | 3337 |
| Flow rate per unit | gpm per unit | - | 84807 | 26696 |
| Brake power per train | hp/train | - | 295 | 52 |
| Brake power per unit | hp/unit | - | 1,181 | 416 |
| *Rich/Lean Amine Exchanger* | | | | |
| Duty per train | MMBtu/h-train | 1,022 | 610 | 150 |
| Duty per unit | MMBtu/h-unit | 4,087 | 2,442 | 1,200 |
| Heat transfer coefficient | Btu/hr-ft2-F | 651 | 651 | 651 |
| LMTD | F | 9.2 | 9.6 | 14.0 |
| Area per train | ft2 | 183,629 | 97,945 | 16,402 |
| Area per unit | ft2 | 734,515 | 391,779 | 131,219 |
| *Rich/Semi-Lean Amine Exchanger* | | | | |
| Duty per train | MMBtu/h-train | - | 93 | 61 |
| Duty per unit | MMBtu/h-unit | - | 373 | 490 |
| Heat transfer coefficient | Btu/hr-ft2-F | - | 651 | 651 |
| LMTD | F | - | 8.2 | 15.4 |
| Area per train | ft2 | - | 17,424 | 6,104 |
| Area per train | ft2 | - | 69,697 | 48,830 |

## Table 4-2. Equipment Comparison Table (English Units, continued)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| *Low Pressure Stripper* | | | | |
| Bottom Pressure | psia | 25 | 25 | 25 |
| Packing type | | | | |
| Diameter | ft | 26.0 | 13.0 | 12.0 |
| Height of packing | ft | 4.9 | 4.9 | 20.0 |
| T-T height | ft | 26 | 26 | n/a |
| Diameter | ft | - | 20.0 | 12.0 |
| Height of packing | ft | - | 12.3 | 25.0 |
| T-T height | ft | - | 40 | 65 |
| *Low Pressure Reboiler* | | | | |
| Number | | | | |
| Duty per train | MMBtu/h-train | 414.2 | 165.2 | 15.0 |
| Duty per unit | MMBtu/h-unit | 1,657 | 661 | 120 |
| Heat transfer coefficient | Btu/h-ft2-F | 150 | 150 | 150 |
| Steam pressure | psia | 35 | 35 | 45 |
| LMTD | F | 37 | 38 | 34 |
| Area | ft2 | 74,312 | 28,789 | 2,982 |
| *Low Pressure Condenser* | | | | |
| Number | | | | |
| Duty per train | MMBtu/h-train | 125 | 33.2 | 16.5 |
| Duty per unit | MMBtu/h-unit | 501 | 133 | 132 |
| Heat transfer coefficient | Btu/h-ft2/F | 80 | 80 | 80 |
| LMTD | F | 50 | 45 | 47 |
| Area per train | ft2 | 31,171 | 9292 | 4352 |
| *Low Pressure Condenser Accumulator* | | | | |
| Number | | | | |
| Diameter | ft | 9.0 | 4 | 6 |
| Length | ft | 38.0 | 16 | 24 |
| *Low Pressure Stripper Condensate Pump* | | | | |
| Flow rate per train | gpm per train | 249 | 57 | 29 |
| Flow rate per unit | gpm per unit | 998 | 228 | 234 |
| Brake power per train | hp/train | 6.7 | 2.0 | 0.8 |
| Brake power per unit | hp/unit | 26.9 | 8 | 6 |
| *Low Pressure Lean Amine Pump* | | | | |
| Flow rate per train | gpm | 28,545 | 20,789 | 27,731 |
| Flow rate per unit | gpm | 114,178 | 83,155 | 3,466 |
| Brake power per train | hp/train | 1,538 | 1,120 | 151 |
| Brake power per unit | hp/unit | 6,153 | 4,481 | 1,205 |
| *Low Pressure Semi-Lean Pump* | | | | |
| Flow rate per train | gpm | - | 4,118 | 1,556 |
| Flow rate per unit | gpm | - | 16,473 | 104 |
| Brake power per train | hp/train | - | 166 | 79 |
| Flow rate per unit | hp/unit | - | 665 | 630 |
| *High Pressure Stripper* | | | | |
| Bottom Pressure | psia | - | 40.5 | 42.9 |
| Packing type | | | | |
| Diameter | ft | - | 22.0 | 15.0 |
| T-T Length | ft | - | 44.0 | 48.0 |
| Height of packing | ft | - | 4.9 | 40.0 |
| *High Pressure Reboiler* | | | | |
| Number | | | | |
| Duty per train | MMBtu/h-train | - | 164 | 100 |
| Duty per unit | MMBtu/h-unit | - | 658 | 800 |
| Heat transfer coefficient | Btu/h-ft2-F | - | 150 | 150 |
| Steam pressure | psia | - | 35 | 45 |
| LMTD | F | - | 40 | 23 |
| Area | ft2 | - | 27,632 | 28,993 |

## Table 4-2. Equipment Comparison Table (English Units, continued)

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| *High Pressure Stripper Condensate Pump* | | | | |
| Flow rate per unit | gpm per unit | - | 124 | 302 |
| Brake power per train | hp/train | - | 2.2 | 1.8 |
| *Lean Cooler* | | | | |
| Duty per train | MMBtu/h-train | 300 | 246 | 35 |
| Duty per unit | MMBtu/h-unit | 1,202 | 985 | 282 |
| Heat transfer coefficient | Btu/h-ft2/F | 140 | 140 | 140 |
| LMTD | F | 21 | 24 | 20 |
| Area | ft2 | 102,690 | 73,081 | 12,813 |
| *Semi-lean Cooler* | | | | |
| Duty per train | MMBtu/h-train | - | 47 | 16 |
| Duty per unit | MMBtu/h-unit | - | 190 | 127 |
| Heat transfer coefficient | Btu/h-ft2/F | - | 140 | 140 |
| LMTD | F | - | 24 | 20 |
| Area | ft2 | - | 14,097 | 5,773 |
| *Lean Surge Tank* | | | | |
| Tank capacity | gal | 412,073 | 300,467 | 48,436 |
| *Makeup Amine Tank* | | | | |
| Tank volume | gal | 117,758 | 121,945 | 118,050 |
| *Makeup Amine Pumps* | | | | |
| Flow rate per unit | gpm per unit | 2.92 | 3.0 | 2.9 |
| *Water Tank* | | | | |
| Capacity | gal | 100,000 | 100,000 | 245,775 |
| *Water Pump* | | | | |
| Flow rate per train | gpm per train | 100 | 100 | 341 |
| | | | | |
| **Equipment Data - CO2 Compression** | | | | |
| *Compressors* | | | | |
| Number of stages | | 4 | 5 | 5 |
| Compressor discharge pressure | psia | 1,400 | 1,400 | 1,400 |
| Total brake power required (total unit) | hp | 54,536 | 51,788 | 47,431 |
| Inlet gas flow rate - Stage 1 | acfm | 98,023 | 75,237 | 36,745 |
| | MMSCFD | 199 | 125 | 70 |
| Inlet gas pressure - Stage 1 | psia | 22.4 | 18 | 21 |
| Inlet gas flow rate - Stage 2 | acfm | 36,270 | 56,285 | 52,225 |
| | MMSCFD | 193 | 195 | 195 |
| Inlet gas pressure - Stage 2 | psia | 57.9 | 38 | 41 |
| | | | | |
| Driver | | steam | steam | |
| Power available from steam | hp | 68,984 | 54,895 | 34,022 |
| Balance of power required from electric driver | hp | | 0 | 13,408 |
| Excess available power | hp | 14,447 | 3,108 | 0 |
| *Compressor Pump (last stage)* | | | | |
| Discharge pressure | psia | 2205 | 2210 | 2205 |
| Total brake power required (total unit) | hp | 2518 | 2520 | 2525 |
| *Compressor Interstage Coolers* | | | | |
| | | | | |
| Type | | | | |
| Total cooler duty | MMBtu/h-unit | 256 | 315 | 381 |
| *Compressor Interstage Separators* | | | | |
| Total capacity | gal | 13,475 | 23,491 | 23,491 |
| *Steam Turbine - CO2 Compressor Driver* | | | | |
| Isentropic efficiency | | | | 72 |
| Inlet temperature | F | 600 | 600 | 600 |
| Inlet pressure | psia | 160 | 160 | 160 |
| Turbine discharge pressure | psia | 35 | 35 | 45 |

**Table 4-2. Equipment Comparison Table (English Units, continued)**

| Description | Units | Base Case | MEA/PZ Double Matrix | MDEA/PZ Double Matrix |
|---|---|---|---|---|
| **Equipment Data - Ancillary Equipment** | | | | |
| *Cooling Water System - Utility* | | | | |
| Water Rates | | | | |
| Stage 1 cooler | gpm | 3973 | 7,399 | 12,941 |
| Stage 2 cooler | gpm | 3681 | 3,172 | 3,069 |
| Stage 3 cooler | gpm | 3605 | 2,981 | 2,908 |
| Stage 4 cooler | gpm | 7389 | 3,129 | 3,082 |
| Stage 5 cooler | gpm | 1853 | 6,733 | 6,665 |
| Stage 6 cooler | gpm | - | 1,849 | 1,855 |
| Lean amine cooler | gpm per train | 24147 | 19,794 | 2,830 |
| Semi-lean amine cooler | gpm per train | - | 3,812 | 1,271 |
| LP stripper condenser | gpm per train | 10059 | 2,672 | 1,327 |
| Total | gpm per unit | 157325 | 130,373 | 73,948 |

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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 commercially available software package that estimates current purchased equipment costs for chemical process equipment. (The costs are in November 2006 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.

- Inlet gas blower – Vendor quote for blower, PDQ$ for motor
- Absorber and Stripper – PDQ$
- Packing for absorber and stripper – Vendor quote
- Pumps (rich/lean/semi-lean, condensate, makeup water and amine, DCC water pump) – PDQ$
- Filtration – PDQ$
- Pressure vessels (reflux accumulator and interstage compression separators) – PDQ$
- Exchangers (reflux condenser, lean amine cooler, reboiler and compressor interstage coolers) – PDQ$
- Amine cross-exchangers (rich/lean exchanger, rich/semi-lean exchanger) – Vendor quote
- 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 5-3 show the major equipment list and purchased equipment costs for the three cases. The major differences in cost stem from the costs of absorbers, amine cross exchangers, and reboilers. The MEA baseline and the MEA / PZ double matrix case use high circulation rates and small $CO_2$ loading changes in the strippers to reduce steam requirements when compared with the previous approaches. This approach transfers costs from the operating domain (derating costs due to reboiler steam requirements) to the capital domain (larger cross exchanger costs). Additionally, the decrease in the steam turbine driver discharge pressure extracts more energy per pound of steam for use in driving the compressor, but this change results in lower temperature steam to the reboilers and thus larger reboilers. Again, this approach has shifted cost from the operating cost domain to the capital cost domain.

The installed costs for purchased equipment (everything but compression) were estimated using a factor methodology similar to that reported in chemical engineering literature (Peters and Timmerhaus, 1991). The installed cost factor for compression was based on vendor recommendations for this type of application. Table 5-4 shows the total process plant cost (PPC) for the different cases. PPC is equivalent to installed equipment cost.

Engineering/home office, project contingency, and process contingency were then added to the total process plant cost to arrive at the total plant cost (TPC). The process plant cost was increased by 10% to account for engineering and home office expenses per the DOE/NETL Systems Analysis Guidelines (SAG) (DOE 2005). 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 (DOE 2005). 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 (DOE 2005; 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 DOE/NETL SAG (DOE 2005), 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;
- Initial catalyst and chemical inventory – 30 days inventory;
- Startup costs – 2% of TPI, 30 days of chemical and operating labor, 7.5 days of fuel (discussed in Section 5.2);
- Spare parts – 0.5% of TPC
- Working capital – 30 days of fuel and consumables, 30 days of byproduct inventory, and 30 days direct expenses (discussed in Section 5.2); and
- Land – 1960 $/acre in 2006 dollars.

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

## Table 5-1. Purchased Equipment Costs for MEA Base Case

| Description | Type | Material of Construction | Design Condition (Per Train Basis) | Number of Trains | Cost per Train | Total Cost |
|---|---|---|---|---|---|---|
| | | | | | | |
| *Inlet Gas Blower* | Centrifugal blower, SS or alloy process-wetter components | SS or alloy process-wetted parts | 2.45 MMkg/h at 10.3 kPa pressure increase (5.40 MMlb/h at 1.5 psi pressure increase) | 1 | 4,350,000 | 4,350,000 |
| *Direct Contact Cooler Water Pump* | Centrifugal, 316SS | 316 SS | 7,050 m3/h at 35 m head (31,100 gpm at 115 ft) | 1 | 456,112 | 456,112 |
| *Absorber* | Packed tower, carbon steel vessel, Diameter = 9.8 m (32 ft), Total packed height = 22.5 m (74 ft), 316SS CMR #2 packing | CS vessel, 316SS packing | 308 kPa and 149 C (300 F and 45 psia) | 4 | 4,870,623 | 19,482,493 |
| *Rich Amine Pump* | Centrifugal, 316SS | 316SS | 6530 m3/h at 52 m head (28,800 gpm at 170 ft) | 4 | 502,095 | 2,008,380 |
| *Rich Amine Carbon Filter* | Carbon steel with teflon gasket | CS with teflon gasket | Treats 15% of rich amine flow | 4 | 150,195 | 600,780 |
| *Particulate Filter* | Carbon steel with teflon gasket | CS with teflon gasket | Treats 15% of rich amine flow | 4 | 150,195 | 600,780 |
| *Rich Amine Booster Pump* | | 316SS | | | | |
| *Rich/Lean Amine Exchanger* | Plate and frame, 316SS rich side, carbon steel lean side | 316SS rich side, CS lean side | Duty = 299 MW (1,020 MMBtu/h) 1140 kPa and 110 C (165 psia and 230 F) | 4 | 7,354,839 | 29,419,355 |
| *Rich/Semi-Lean Amine Exchanger* | | 316SS rich side, CS semi-lean side | | n/a | n/a | n/a |
| *LP Stripper - Upper Section* | Packed tower, carbon steel vessel, Diameter = 7.9 m (26 ft), Total packed height = 1.5 m (5 ft), 316SS CMR#2 packing | CS vessel, 316SS packing | 260 kPa and 149 C (38 psia and 300 F) | 4 | 551,681 | 2,206,723 |
| *LP Stripper - Lower Section* | | CS vessel, 316SS packing | | | | |
| *LP Reboiler* | Kettle-type shell and tube, 316SS tubes and carbon steel shell | 316SS tubes and CS shell | Duty = 121 MW (414 MMBtu/h) 446 kPa and 177 C (65 psia and 350 F) | 4 | 3,882,142 | 15,528,568 |
| *LP Condenser* | Shell and tube, 316SS tubes and carbon steel shell | 316SS tubes and CS shell | Duty = 36.7 MW (125 MMBtu/h) 446 kPa and 177 C (65 psia and 350 F) | 4 | 925,456 | 3,701,824 |
| *LP Condenser Accumulator* | Horizontal 316SS vessel | 316SS | 446 kPa and 177 C (65 psia and 350 F) | 4 | 113,893 | 455,572 |
| *LP Condensate Pump* | Centrifugal, 316SS | 316SS | 57 m3/h at 21 m head (249 gpm at 70 ft) | 4 | 11,064 | 44,256 |
| *Lean Amine Pump* | Centrifugal, 316SS | 316SS | 6480 m3/h at 45 m head (28,500 gpm at 148 ft) | 4 | 367,980 | 1,471,920 |
| *Semi-Lean Amine Pump* | | 316SS | | n/a | n/a | n/a |

## Table 5-1. Purchased Equipment Costs for MEA Base Case (continued)

| Description | Type | Material of Construction | Design Condition (Per Train Basis) | Number of Trains | Cost per Train | Total Cost |
|---|---|---|---|---|---|---|
| *HP Stripper* | | CS vessel, 316SS packing | | n/a | n/a | n/a |
| *HP Reboiler* | | 316SS tubes and CS shell | | n/a | n/a | n/a |
| *HP Condenser* | | | | n/a | n/a | n/a |
| *HP Condenser Accumulator* | | | | n/a | n/a | n/a |
| *HP Condensate Pump* | | 316SS | | n/a | n/a | n/a |
| *Lean Cooler* | Shell and tube, 316SS tubes and carbon steel shell, water cooled | 316SS tubes and CS shell | Duty = 88.0 MW (300 MMBtu/h) 791 kPa and 177 C (115 psia and 350 F) | 4 | 2,679,761 | 10,719,044 |
| *Semi-Lean Cooler* | | 316SS tubes and CS shell | | n/a | n/a | n/a |
| *Compressors* | Centrifugal, four stages (plus pump, separate line item), 316SS | 316SS | Discharge pressure = 9650 kPa, (1400 psia) | 1 | 9,346,310 | 9,346,310 |
| *Compressor Driver - steam turbine* | | | Discharge pressure = 239 kPa (34.7 psia) | 1 | 12,300,000 | 12,300,000 |
| *Compressor Driver - electric* | | | | n/a | n/a | n/a |
| *Compressor Pump (last stage)* | Electric submersible Pump (ESP), 316SS | 316SS | Discharge pressure = 15,200 kPa (2205 psia) | 1 | 669,786 | 669,786 |
| *Compressor Interstage Coolers* | Shell and tube, 316SS tubes and carbon steel shell, water cooled | 316SS tubes and CS shell | Total duty = 75 MW (therm) (256 MMBtu/h) | 1 | 2,148,968 | 2,148,968 |
| *Compressor Interstage Separators* | 316SS horizontal vessels | 316SS | 5 minute residence time | 1 | 222,537 | 222,537 |
| *Makeup Amine Tank* | Fixed roof tank | CS | 446 m3 (118,000 gal) | 1 | 89,871 | 89,871 |
| *Makeup Amine Pumps* | Centrifugal | CS | 4.5 m3/h at 37 m head (20 gpm at 120 ft head) | 1 | 5,406 | 5,406 |
| *Water Tank* | Fixed roof tank | CS | 379 m3 (100,000 gal) | 1 | 89,871 | 89,871 |
| *Water Pump* | Centrifugal | CS | 23 m3/h at 31 m head (100 gpm at 100 ft head) | 1 | 8,000 | 8,000 |
| *Surge Tank* | Fixed roof tank | CS | 1,560 m3 (412,000 gal) | 4 | 400,892 | 1,603,568 |
| *Reclaimer* | | | | 4 | 2,774,916 | 11,099,662 |
| *Cooling Water System - DCC* | Includes cooling tower, basin, fans, and pumps | | 7,050 m3/h (31,100 gpm) | 1 | 1,227,690 | 1,227,690 |
| *Cooling Water System - Utility* | Includes cooling tower, basin, fans, and pumps | | 35,700 m3/h (157,000 gpm) | 1 | 11,613,560 | 11,613,560 |
| *Dehydration Unit* | | | 225,000 std m3/h (190 MMSCFD) | 1 | 1,557,110 | 1,557,110 |
| | | | | | | 143,028,147 |

# Table 5-2. Purchased Equipment Costs for MEA / PZ Double Matrix

| Description | Type | Materials of Construction | Design Condition (Per Train Basis) | Number of Trains | Cost Per Train | Total Cost |
|---|---|---|---|---|---|---|
| | | | | | | |
| *Inlet Gas Blower* | Centrifugal blower | SS or alloy process-wetted parts | 2.45 MMkg/h at 10.3 kPa pressure increase (5.40 MMlb/h at 1.5 psi pressure increase) | 1 | 4,350,000 | 4,350,000 |
| *Direct Contact Cooler Water Pump* | Centrifugal | 316 SS | 7,050 m3/h at 35 m head (31,100 gpm at 115 ft) | 1 | 456,112 | 456,112 |
| *Absorber* | Packed tower, Diameter = 10.7 m (35 ft), Total packed height = 22.5 m (74 ft), CMR #2 packing | CS vessel, 316SS packing | 308 kPa and 149 C (300 F and 45 psia) | 4 | 5,758,980 | 23,035,919 |
| *Rich Amine Pump* | Centrifugal | 316SS | 5730 m3/h at 53 m head (25,200 gpm at 173 ft) | 4 | 435,149 | 1,740,596 |
| *Rich Amine Carbon Filter* | | CS with teflon gasket | Treats 15% of rich amine flow | 4 | 145,902 | 583,608 |
| *Particulate Filter* | | CS with teflon gasket | Treats 15% of rich amine flow | 4 | 145,902 | 583,608 |
| *Rich Amine Booster Pump* | Centrifugal | 316SS | 4820 m3/h at 12 m head (21,200 gpm at 38 ft) | 4 | 259,932 | 1,039,728 |
| *Rich/Lean Amine Exchanger* | Plate and frame | 316SS rich side, CS lean side | Duty = 179 MW (610 MMBtu/h) 1140 kPa and 110 C (165 psia and 230 F) | 4 | 4,179,743 | 16,718,971 |
| *Rich/Semi-Lean Amine Exchanger* | Plate and frame | 316SS rich side, CS semi-lean side | Duty = 27.3 MW (93 MMBtu/h) 1140 kPa and 110 C (165 psia and 230 F) | 4 | 779,227 | 3,116,906 |
| *LP Stripper - Upper Section* | Packed tower, Diameter = 4 m (13 ft), Total packed height = 1.5 m (5 ft), CMR#2 packing | CS vessel, 316SS packing | 260 kPa and 149 C (38 psia and 300 F) | 4 | 197,340 | 789,359 |
| *LP Stripper - Lower Section* | Packed tower, Diameter = 6.1 m (20 ft), Total packed height = 3.8 m (12 ft), CMR#2 packing | CS vessel, 316SS packing | 260 kPa and 149 C (38 psia and 300 F) | 4 | 552,338 | 2,209,354 |
| *LP Reboiler* | Kettle-type shell and tube | 316SS tubes and CS shell | Duty = 48.4 MW (165 MMBtu/h) 446 kPa and 177 C (65 psia and 350 F) | 4 | 1,573,520 | 6,294,080 |
| *LP Condenser* | Shell and tube | 316SS tubes and CS shell | Duty = 9.7 MW (33 MMBtu/h) 446 kPa and 177 C (65 psia and 350 F) | 4 | 338,790 | 1,355,160 |
| *LP Condenser Accumulator* | Horizontal vessel | 316SS | 446 kPa and 177 C (65 psia and 350 F) | 4 | 23,049 | 92,196 |
| *LP Condensate Pump* | Centrifugal | 316SS | 13 m3/h at 28 m head (57 gpm at 93 ft) | 4 | 8,924 | 35,696 |
| *Lean Amine Pump* | Centrifugal | 316SS | 4,720 m3/h at 46 m head (20,800 gpm at 148 ft) | 4 | 306,650 | 1,226,600 |
| *Semi-Lean Amine Pump* | Centrifugal | 316SS | 681 m3/h at 36 m head (3,000 gpm at 118 ft) | 4 | 78,588 | 314,352 |

## Table 5-2. Purchased Equipment Costs for MEA / PZ Double Matrix (continued)

| Description | Type | Materials of Construction | Design Condition (Per Train Basis) | Number of Trains | Cost Per Train | Total Cost |
|---|---|---|---|---|---|---|
| *HP Stripper* | Packed tower, Diameter = 6.7 m (22 ft), Total packed height = 1.5 m (5 ft), CMR#2 packing | CS vessel, 316SS packing | 446 kPa and 177 C (65 psia and 350 F) | 4 | 512,565 | 2,050,261 |
| *HP Reboiler* | Kettle-type shell and tube | 316SS tubes and CS shell | Duty = 48.2 MW (164 MMBtu/h)<br>446 kPa and 177 C (65 psia and 350 F) | 4 | 1,519,425 | 6,077,700 |
| *HP Condensate Pump* | Centrifugal | 316SS | 28 m3/h at 14 m head (124 gpm at 47 ft) | 1 | 8,088 | 8,088 |
| *Lean Cooler* | Shell and tube, water cooled | 316SS tubes and CS shell | Duty = 72.2 MW (246 MMBtu/h)<br>791 kPa and 177 C (115 psia and 350 F) | 4 | 1,921,255 | 7,685,020 |
| *Semi-Lean Cooler* | Shell and tube, water cooled | 316SS tubes and CS shell | Duty = 13.9 MW (47.4 MMBtu/h)<br>791 kPa and 177 C (115 psia and 350 F) | 4 | 474,177 | 1,896,708 |
| *Compressors* | Centrifugal, five stages (plus pump, separate line item) | 316SS | Discharge pressure = 9650 kPa, (1400 psia) | 1 | 8,943,740 | 8,943,740 |
| *Compressor Driver - steam turbine* | | | Discharge pressure = 239 kPa (34.7 psia) | 1 | 10,724,660 | 10,724,660 |
| *Compressor Driver - electric* | | | | n/a | n/a | n/a |
| *Compressor Pump (last stage)* | Electric submersible Pump (ESP) | 316SS | Discharge pressure = 15,200 kPa (2205 psia) | 1 | 669,786 | 669,786 |
| *Compressor Interstage Coolers* | Shell and tube, water cooled | 316SS tubes and CS shell | Total duty = 92 MW (therm) (315 MMBtu/h) | 1 | 3,608,784 | 3,608,784 |
| *Compressor Interstage Separators* | Horizontal vessels | 316SS | 5 minute residence time | 1 | 493,786 | 493,786 |
| *Makeup Amine Tank* | Fixed roof tank | CS | 462 m3 (122,000 gal) | 1 | 89,871 | 89,871 |
| *Makeup Amine Pumps* | Centrifugal | CS | 4.5 m3/h at 37 m head (20 gpm at 120 ft head) | 1 | 5,406 | 5,406 |
| *Water Tank* | Fixed roof tank | CS | 379 m3 (100,000 gal) | 1 | 89,871 | 89,871 |
| *Water Pump* | Centrifugal | CS | 23 m3/h at 31 m head (100 gpm at 100 ft head) | 1 | 8,000 | 8,000 |
| *Surge Tank* | Fixed roof tank | CS | 1,140 m3 (300,000 gal) | 4 | 147,548 | 590,192 |
| *Reclaimer* | | | | 4 | 2,774,916 | 11,099,662 |
| *Cooling Water System - DCC* | Includes cooling tower, basin, fans, and pumps | | 7,050 m3/h (31,100 gpm) | 1 | 1,227,690 | 1,227,690 |
| *Cooling Water System - Utility* | Includes cooling tower, basin, fans, and pumps | | 29,600 m3/h (130,000 gpm) | 1 | 9,443,780 | 9,443,780 |
| *Dehydration Unit* | | | 225,000 std m3/h (190 MMSCFD) | 1 | 1,558,732 | 1,558,732 |
| ***Total*** | | | | | | 130,213,982 |

# Table 5-3. Purchased Equipment Costs for MDEA / PZ Double Matrix

| Description | Type | Materials of Construction | Design Condition (Per Train Basis) | Number of Trains | Cost per Train | Total Cost |
|---|---|---|---|---|---|---|
| *Inlet Gas Blower* | Centrifugal blower | SS or alloy process-wetted parts | 2.45 MMkg/h at 10.3 kPa pressure increase (5.40 MMlb/h at 1.5 psi pressure increase) | 1 | 4,350,000 | 4,350,000 |
| *Direct Contact Cooler Water Pump* | Centrifugal | 316 SS | 7,050 m3/h at 35 m head (31,100 gpm at 115 ft) | 1 | 456,112 | 456,112 |
| *Absorber* | Packed tower, Diameter = 11.9 m (39 ft), Total packed height = 15 m (50 ft), Flexipac 1Y | CS vessel, 316SS packing | 308 kPa and 149 C (300 F and 45 psia) | 8 | 5,080,455 | 40,643,639 |
| *Rich Amine Pump* | Centrifugal | 316SS | 1,060 m3/h at 45 m head (4,650 gpm at 147 ft of head) | 8 | 91,995 | 735,960 |
| *Rich Amine Carbon Filter* | | CS with teflon gasket | Treats 15% of rich amine flow | 8 | 25,684 | 205,472 |
| *Particulate Filter* | | CS with teflon gasket | Treats 15% of rich amine flow | 8 | 25,684 | 205,472 |
| *Rich Amine Booster Pump* | Centrifugal | 316SS | 758 m3/h at 12 m head (3,340 gpm at 38 ft of head) | 8 | 49,920 | 399,360 |
| *Rich/Lean Amine Exchanger* | Plate and frame | 316SS rich side, CS lean side | Duty = 44 MW (150 MMBtu/h) 1140 kPa and 110 C (165 psia and 230 F) | 8 | 751,478 | 6,011,826 |
| *Rich/Semi-Lean Amine Exchanger* | Plate and frame | 316SS rich side, CS semi-lean side | Duty = 18 MW (61 MMBtu/h) 1140 kPa and 110 C (165 psia and 230 F) | 8 | 415,273 | 3,322,180 |
| *LP Stripper - Upper Section* | Packed tower, Diameter = 3.7 m (12 ft), Total packed height = 6.1 m (20 ft), Flexipac 1Y packing | CS vessel, 316SS packing | 260 kPa and 149 C (38 psia and 300 F) | 8 | 158,336 | 1,266,690 |
| *LP Stripper - Lower Section* | Packed tower, Diameter = 3.7 m (12 ft), Total packed height = 7.6 m (25 ft), Flexipac 1Y packing | CS vessel, 316SS packing | 260 kPa and 149 C (38 psia and 300 F) | 8 | 371,391 | 2,971,131 |
| *LP Reboiler* | Kettle-type shell and tube | 316SS tubes and CS shell | 4.4 MW (15 MMBtu/h), 791 kPa and 204 C (115 psia and 400 F) | 8 | 242,637 | 1,941,096 |
| *LP Condenser* | Shell and tube | 316SS tubes and CS shell | 4.8 MW (16.5 MMBtu/h), 446 kPa and 149 C (65 psia and 300 F) | 8 | 209,091 | 1,672,728 |
| *LP Condenser Accumulator* | Horizontal vessel | 316SS | 4.8 MW (16.5 MMBtu/h), 446 kPa and 149 C (65 psia and 300 F) | 8 | 70,764 | 566,112 |
| *LP Condensate Pump* | Centrifugal | 316SS | 7m3/h at 21 m of head (29 gpm at 70 ft of head) | 8 | 8,088 | 64,704 |
| *Lean Amine Pump* | Centrifugal | 316SS | 787 m3/h at 36 m of head (27,700 gpm at 119 ft of head) | 8 | 71,778 | 574,224 |
| *Semi-Lean Amine Pump* | Centrifugal | 316SS | 353 m3/h at 32 m of head (3,470 gpm at 104 ft of head) | 8 | 42,214 | 337,712 |

## Table 5-3. Purchased Equipment Costs for MDEA / PZ Double Matrix (continued)

| Description | Type | Materials of Construction | Design Condition (Per Train Basis) | Number of Trains | Cost per Train | Total Cost |
|---|---|---|---|---|---|---|
| *HP Stripper* | Packed tower, Diameter = 4.6 m (15 ft), Total packed height = 12 m (40 ft), Flexipac 1Y packing | CS vessel, 316SS packing | 446 kPa and 149 C (65 psia and 300 F) | 8 | 712,284 | 5,698,271 |
| *HP Reboiler* | Kettle-type shell and tube | 316SS tubes and CS shell | 29 MW (100 MMBtu/h), 791 kPa and 204 C (115 psia and 400 F) | 8 | 1,597,095 | 12,776,760 |
| *HP Condensate Pump* | Centrifugal | 316SS | 69 m3/h at 8 m of head (302 gpm at 27 ft of head) | 1 | 11,680 | 11,680 |
| *Lean Cooler* | Shell and tube, water cooled | 316SS tubes and CS shell | 10.3 MW (35.2 MMBtu/h) 791 kPa and 177 C (115 psia and 350 F) | 8 | 442,628 | 3,541,024 |
| *Semi-Lean Cooler* | Shell and tube, water cooled | 316SS tubes and CS shell | 4.6 MW (15.8 MMBtu/h) 791 kPa and 177 C (115 psia and 350 F) | 8 | 257,498 | 2,059,984 |
| *Compressors* | Centrifugal, five stages (plus pump, separate line item) | 316SS | Discharge pressure = 9650 kPa, (1400 psia) | 1 | 6,938,700 | 6,938,700 |
| *Compressor Driver - steam turbine* | | | | 1 | 8,701,782 | 8,701,782 |
| *Compressor Driver - electric* | | | | 1 | 7,800,000 | 7,800,000 |
| *Compressor Pump (last stage)* | Electric submersible Pump (ESP) | 316SS | Discharge pressure = 15,200 kPa (2205 psia) | 1 | 670,979 | 670,979 |
| *Compressor Interstage Coolers* | Shell and tube, water cooled | 316SS tubes and CS shell | Total duty = 112 MW (therm) (381 MMBtu/h) | 1 | 4,483,757 | 4,483,757 |
| *Compressor Interstage Separators* | Horizontal vessels | 316SS | 5 minute residence time | 1 | 493,786 | 493,786 |
| *Makeup Amine Tank* | Fixed roof tank | CS | 447 m3 (118,000 gal) | 1 | 89,871 | 89,871 |
| *Makeup Amine Pumps* | Centrifugal | CS | 4.5 m3/h at 37 m head (20 gpm at 120 ft head) | 1 | 5,406 | 5,406 |
| *Water Tank* | Fixed roof tank | CS | 930 m3 (246,000 gal) | 1 | 127,094 | 127,094 |
| *Water Pump* | Centrifugal | CS | 78 m3/h (341 gpm) | 1 | 31,936 | 31,936 |
| *Surge Tank* | Fixed roof tank | CS | 183 m3 (48,400 gal) | 8 | 64,121 | 512,968 |
| *Reclaimer* | | | | 8 | 1,387,458 | 11,099,662 |
| *Cooling Water System - DCC* | Includes cooling tower, basin, fans, and pumps | | 7,050 m3/h (31,100 gpm) | 1 | 1,227,690 | 1,227,690 |
| *Cooling Water System - Utility* | Includes cooling tower, basin, fans, and pumps | | 16,800 m3/h (73,900 gpm) | 1 | 5,060,320 | 5,060,320 |
| *Dehydration Unit* | | | | 1 | 1,560,417 | 1,560,417 |
| ***Total*** | | | | | | 138,616,505 |

**Table 5-4. Process Plant Costs**

| | Factors | | Costs | | | |
|---|---|---|---|---|---|---|
| **Description** | **Units** | **Value** | **Units** | **MEA Base Case** | **MEA / PZ Double Matrix** | **MDEA / PZ Double Matrix** |
| **Process Plant Cost (PPC)** | | | | | | |
| CO2 capture purchased equipment costs | | | $ | 118,341,000 | 105,773,000 | 109,528,000 |
| Installation | | | | | | |
| Total installation | % of capture PEC | 104 | $ | 123,074,000 | 132,217,000 | 136,909,000 |
| CO2 capture installed equipment costs | | | $ | 241,415,000 | 237,990,000 | 246,437,000 |
| CO2 compression purchased equipment costs | | | $ | 24,688,000 | 24,441,000 | 29,089,000 |
| Installation | % of compression PEC | 180 | $ | 44,438,000 | 43,993,000 | 52,360,000 |
| CO2 compression installed equipment costs | | | $ | 69,125,000 | 68,434,000 | 81,449,000 |
| Process Plant Cost (PPC) | | | $ | 310,540,000 | 306,424,000 | 327,886,000 |

# Table 5-5. Total Capital Requirement

| | Factors | | Costs | | | |
|---|---|---|---|---|---|---|
| Description | Units | Value | Units | MEA Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
| **Process Plant Cost (PPC)** | | | $ | 310,540,000 | 306,424,000 | 327,886,000 |
| **Total Plant Cost (TPC)** | | | | | | |
| Engineering cost | % of PPC | 10 | $ | 31,054,000 | 30,642,000 | 32,789,000 |
| Process Contingency | % of PPC | 5 | $ | 15,527,000 | 15,321,000 | 16,394,000 |
| Project Contingency | % of PPC | 30 | $ | 93,162,000 | 91,927,000 | 98,366,000 |
| Total Plant Cost (TPC) | | | $ | 450,283,000 | 444,315,000 | 475,435,000 |
| **Total Plant Investment (TPI)** | | | | | | |
| Interest and Inflation Factor | % of PPC | 10 | $ | 31,054,000 | 30,642,000 | 32,789,000 |
| Total Plant Investment | | | $ | 481,337,000 | 474,957,000 | 508,223,000 |
| **Total Capital Requirement (TCR)** | | | | | | |
| Prepaid royalties | % of PPC | 0.5 | $ | 1,553,000 | 1,532,000 | 1,639,000 |
| Initial catalyst and chemical inventory | days inventory | 30 | $ | 557,000 | 571,000 | 600,000 |
| Startup Costs | | | | | | |
| Component 1 | % of TPI | 2 | $ | 9,627,000 | 9,499,000 | 10,164,000 |
| Component 2 | days chemicals and operating labor | 30 | $ | 641,000 | 623,000 | 599,000 |
| Component 3 | days fuel inventory | 7.5 | $ | | | |
| Spare parts | % of TPC | 0.5 | $ | 2,251,000 | 2,222,000 | 2,377,000 |
| Working Capital | | | | | | |
| Component 1 | days fuel and consumables | 30 | $ | 634,000 | 616,000 | 592,000 |
| Component 2 | days byproduct inventory | 30 | $ | | | |
| Component 3 | days direct expenses | 30 | $ | 871,000 | 859,000 | 919,000 |
| Land | 1960 | $/acre | | | | |
| | 20 | acres | $ | 39,000 | 39,000 | 39,000 |
| Total Capital Requirement | | | $ | 497,509,000 | 490,918,000 | 525,154,000 |

## 5.2 Operating Costs

The major operating and maintenance (O&M) costs for the $CO_2$ capture and compression process include consumables, maintenance costs, plant labor, and byproduct credits as shown in Table 5-6. This process does not have byproduct credits. 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 5-6. Operating and Maintenance Cost Parameters and Values**

| **Consumables O&M Cost Components** | **Value** | **Units** |
|---|---|---|
| Solvent cost | 1200 | $/tonne |
| Water cost | 0.92 | $/1000 gal |
| Solid waste disposal cost | 175 | $/tonne |
| **Maintenance Costs** | | |
| Total maintenance cost | 2.2 | % of total plant cost |
| **Labor Costs** | | |
| Maintenance cost allocated to labor | 12 | % of maintenance costs |
| Administration and support labor cost | 30 | % of operating labor |
| Operating labor | 1 | # operators |

The O&M cost factors, except for operating labor, were obtained from the Systems Analysis Guidelines (DOE 2005). 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 80% was used as recommended in the Systems Analysis Guidelines.

The variable O&M components include costs of chemicals consumed, utilities, and services used. Solvent 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 solvent for electric utility type operations. Quantifying solvent degradation losses is still a major technical unknown for these systems. Therefore, similar degradation rates and solvent unit costs were assumed for all solvents. 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 amine. 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 O&M costs are shown in Table 5-7.

## Table 5-7. Summary of Operating and Maintenance Costs

| Description | Units | MEA Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| **Total Operating Costs (TOC)** | | | | |
| Consumables (1 year at capacity factor) | | | | |
| Chemicals | | | | |
| Solvent unit cost | $/tonne | 1,200 | | |
| Solvent usage | kg/h | 644 | 661 | 695 |
| Solvent cost per year - CO2 captur | $/year | 5,417,000 | 5,556,000 | 5,841,000 |
| Water | | | | |
| Water unit cost | $/1000 gal | 0.92 | = | = |
| Makeup water rate | gpm | 5845 | 4915 | 3433 |
| Water cost per year - CO2 capture | $/year | 2,261,000 | 1,901,000 | 1,328,000 |
| Mercury removal (activated carbon) | $/year | 0 | = | = |
| Waste disposal | | | | |
| | $/tonne | 175 | = | = |
| | kg waste/ tonne CO2 captured | 0.075 | = | = |
| | tonne/year | 218 | = | = |
| | $/year | 38,000 | 38,000 | 38,000 |
| Fuel | $/year | 0 | 0 | 0 |
| Total consumables - CO2 Capture | $/year | 7,717,000 | 7,496,000 | 7,207,000 |
| Maintenance Costs | | | | |
| Factor | % of TPC | 2.2 | = | = |
| Total maintenance costs | $/year | 9,906,000 | 9,775,000 | 10,460,000 |
| Plant Labor | | | | |
| Operating labor | 1 operator/year | 80,000 | 80,000 | 80,000 |
| Supervisory and clerical labor | | | | |
| Component 1 | % of operating labor | 30 | = | = |
| | $/year | 24,000 | 24,000 | 24,000 |
| Component 2 | % of maintenance cost | 12 | | |
| | $/year | 1,189,000 | 1,173,000 | 1,255,000 |
| Total labor costs | $/year | 1,293,000 | 1,277,000 | 1,359,000 |
| Total Operating Costs | $/year | 18,916,000 | 18,548,000 | 19,026,000 |
| **Byproduct Credits** | | | | |
| Unit price for CO2 | $/tonne | 0 | = | = |
| Total byproduct credits | $/year | 0 | 0 | 0 |
| **Net Operating Cost - CO2 Capture** | $/year | 18,916,000 | 18,548,000 | 19,026,000 |

*"=" indicates equal to the base case.*

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 next section describes the derating calculations.

### 5.3 Derating

The energy requirements to operate the main facility and the $CO_2$ capture unit are withdrawn from the main power facility output either through electricity or steam. These energy draws decrease the net electrical output of the plant; they derate the plant. Power requirements of electric motors translate directly to electrical derating (a decrease in MWe). Energy requirements that are supplied using steam, such as the heat requirements for the reboilers, must be converted into electrical derating by calculating the amount of electrical generating capacity that the steam would have supplied to the main power facility had the steam not been diverted to the $CO_2$ capture system. The derating factor used for this study was 70 W derating for every 1 lb/h of steam diverted from the LP turbine of the main facility (145 W/(kg/h) ). Table 5-8 summarizes the energy requirements for the three cases and then shows the electrical derating that results from each of the energy requirements. Finally, Table 5-8 shows the net generating capacity for the 500 MWe gross plant with and without $CO_2$ capture. Table 5-9 lists how the various energy demands correlate to electrical derating.

The MEA base case and the MEA / PZ double matrix case both use 20 psig steam (239 kPa, 34.7 psia). The energy provided by the $CO_2$ capture steam turbine exceeded the energy requirements to drive the $CO_2$ compressors. This energy was added back to the net generating capacity. The MDEA / PZ double matrix case required 30 psig steam (308 kPa, 44.7 psia). The energy supplied by the $CO_2$ capture steam turbine was not enough to drive the compressors; therefore, a supplemental electrical motor was included for this case.

The base plant has auxiliary power requirements of ~29 MW, so the net capacity without capture is 471 MWe. With $CO_2$ capture, the MEA base case net capacity is 339 MWe, the MEA / PZ double matrix net generating capacity is 356 MWe, and the MDEA / PZ double matrix is 378 MWe. Thus, when compared to the MEA base case, the MEA / PZ double matrix decreases the derating due to $CO_2$ capture by an estimated 13% (from 132.0 MWe to 115.2 MWe), and the MDEA / PZ double matrix decreases the derating due to $CO_2$ capture by an estimated 30% (from 132.0 MWe to 93.0 MWe).

## Table 5-8. Derating Results

| Description | Units | MEA Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| *Base Plant Gross Capacity* | MWe | 500.0 | 500.0 | 500.0 |
| *Pumps and Compressors* Note 1 | | | | |
| Inlet gas blower | MW | 8.4 | 8.4 | 8.4 |
| Direct contact water cooler pump | MW | 1.0 | 1.0 | 1.0 |
| Rich amine pump | MW | 5.4 | 4.7 | 1.7 |
| Rich amine booster pump | MW | - | 0.9 | 0.3 |
| LP lean amine pump | MW | 4.6 | 3.3 | 0.8 |
| LP semi-lean pump | MW | - | 0.5 | 0.9 |
| HP lean pump | MW | - | - | - |
| Cooling tower pumps | MW | 3.7 | 3.1 | 1.9 |
| Compression - 1st Stage | MW | 10.7 | 4.9 | 2.5 |
| Compression - 2nd Stage | MW | 11.2 | 9.3 | 9.1 |
| Compression - 3rd Stage | MW | 10.1 | 9.1 | 8.8 |
| Compression - 4th Stage | MW | 8.7 | 8.2 | 8.0 |
| Compression - 5th Stage | MW | - | 7.1 | 6.9 |
| Compression - ESP | MW | 1.9 | 1.9 | 1.9 |
| *Heat Exchangers* Note 1 | | | | |
| Lean cooler | MW | 352.2 | 288.7 | 82.6 |
| Semi-lean cooler | MW | - | 55.6 | 37.1 |
| Rich-lean exchanger | MW | 1197.8 | 715.6 | 351.7 |
| Rich/semi-lean exchanger | MW | - | 109.2 | 143.5 |
| LP stripper condenser | MW | 146.7 | 39.0 | 38.7 |
| Stg 1 cooler | MW | 14.5 | 27.1 | 47.3 |
| Stg 2 cooler | MW | 13.5 | 11.6 | 11.2 |
| Stg 3 cooler | MW | 13.1 | 10.9 | 10.6 |
| Stg 4 cooler | MW | 27.0 | 11.4 | 11.3 |
| Stg 5 cooler | MW | | 24.6 | 24.4 |
| ESP cooler | MW | 6.8 | 6.8 | 6.8 |
| HP reboiler | MW | | 192.7 | 234.5 |
| LP reboiler | MW | 485.5 | 193.7 | 35.2 |
| *Subtotals* Note 1 | | | | |
| Inlet blower work | MW | 8.4 | 8.4 | 8.4 |
| Total pump work | MW | 14.7 | 13.6 | 6.6 |
| Total compressor work | MW | 40.7 | 38.6 | 35.4 |
| CO2 pump work | MW | 1.9 | 1.9 | 1.9 |
| Total cooling water duty | MW | 573.9 | 475.7 | 270.0 |
| Total reboiler heat duty | MW | 485.5 | 386.4 | 269.7 |
| *Auxiliary Loads* | | | | |
| Base plant, PC | MWe | 29.1 | 29.1 | 29.1 |
| FGD system | MWe | 0.0 | 0.0 | 0.0 |
| ESP | MWe | 0.0 | 0.0 | 0.0 |
| SCR | MWe | 0.0 | 0.0 | 0.0 |
| *Main facility derating* | *MWe* | *29.1* | *29.1* | *29.1* |
| Inlet blower work (direct electricity usage) | MWe | 8.4 | 8.4 | 8.4 |
| Pump work (direct electricity usage) | MWe | 14.7 | 13.6 | 6.6 |
| Reboiler heat (steam derating) | MWe | 117.8 | 93.7 | 66.2 |
| Compressor work (supplemental electric driver) | MWe | 0.0 | 0.0 | 10.0 |
| CO2 pump work | MWe | 1.9 | 1.9 | 1.9 |
| Excess energy from turbine | MWe | -10.8 | -2.3 | 0.0 |
| *CO2 capture derating* | *MWe* | *132.0* | *115.2* | *93.0* |
| *Total derating* | *MWe* | *161.0* | *144.3* | *122.1* |
| **Plant Net Electrical Capacity - without capture** | **MWe** | **471** | **471** | **471** |
| **Plant Net Electrical Capacity - with capture** | **MWe** | **339** | **356** | **378** |

*Note 1: These MW values are not all directly related to MWe derating. See Table 5-9.*

**Table 5-9. Effect of Energy Requirements on Derating**

| Equipment | Type of Energy | Effect on Derating |
|---|---|---|
| Inlet blower work | Electricity | Directly related to derate |
| Total pump work | Electricity | Directly related to derate |
| Total compressor work | Steam driver | Must be converted |
| CO2 pump work | Electricity | Directly related to derate |
| Total cooling water duty | Cooling water | Does not derate |
| Total reboiler heat duty | Steam | Must be converted |

## 5.4 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 14% is used in the analysis for the cases as recommended by the DOE/NETL SAG (DOE 2005). Table 5-10 shows how these parameters vary for the different cases. The MEA base case and MEA / PZ double matrix only differ by a few percent, and the MDEA / PZ double matrix annual revenue requirement is less than 5% greater than the baseline. However, as will be discussed in Section 6, the differences in net generating capacity will affect the final economic analysis based on cost of electricity and cost of avoided $CO_2$ emissions.

**Table 5-10. Total Annual Revenue Requirement**

| Description | Units | MEA Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| Levelized capital charge factor | % / year | 14 | 14 | 14 |
| Annual $CO_2$ capture capital costs | MM$/yr | 69.7 | 68.7 | 73.5 |
| Annual $CO_2$ capture operating costs | MM$/yr | 18.9 | 18.5 | 19.0 |
| Total Annual $CO_2$ capture Revenue Requirement | MM$/yr | 88.6 | 87.3 | 92.5 |

**References (Section 5)**

Department of Energy (DOE) National Energy Technology Laboratory (NETL). "Carbon Capture and Sequestration Systems Analysis Guidelines", April 2005.

*Evaluation of Innovative Fossil Fuel Power Plants with $CO_2$ Removal*, EPRI, Palo Alto, CA, U.S. Department of Energy – Office of Fossil Energy, Germantown, MD and U.S. Department of Energy/NETL, Pittsburgh, PA: 2000. 1000316.

Peters and Timmerhaus. *Plant Design and Economics for Chemical Engineers*, 4th Edition, Chapter 5, 1991.

Rao, Anand; E. Rubin; M. Berkenpas; "*An Integrated Modeling Framework for Carbon Management Technologies*", U.S. Department of Energy, March 2004.

# 6.0 ECONOMIC ANALYSIS AND RESULTS

This section utilizes the annualized cost summary from Section 5 to compare the cost of electricity and the cost of $CO_2$ avoidance for the three cases.

## 6.1 Cost of Electricity

Table 6-1 presents the cost of electricity with and without $CO_2$ capture for the three cases. The base plant cost of electricity is assumed as 5 cents/kWh. The basis for these costs was previously presented in Section 5.

**Table 6-1. Cost of Electricity**

| Description | Units | MEA Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|
| Gross generating capacity | MWe | 500 | 500 | 500 |
| Net generating capacity without $CO_2$ capture | MWe | 471 | 471 | 471 |
| Net generating capacity with $CO_2$ capture | MWe | 339 | 356 | 378 |
| Base plant cost of electricity | c/kWh | 5.0 | 5.0 | 5.0 |
| Annual base plant costs | MM$/yr | 165.0 | 165.0 | 165.0 |
| Total annual $CO_2$ capture costs | MM$/yr | 88.6 | 87.3 | 92.5 |
| Total annual costs with $CO_2$ capture | MM$/yr | 253.6 | 252.3 | 257.6 |
| Total COE | c/kWh | 10.7 | 10.1 | 9.7 |
| Increase in COE | % | 113 | 102 | 95 |

As shown in the table, the cost of electricity is highest (10.7 c/kWh) for the MEA base case. The increase in cost of electricity for the base case is estimated at 113%. For the MEA / PZ double matrix, the estimated increase is 102%. Finally, the estimated increase in cost of electricity for the MDEA / PZ double matrix is 95%, which is the smallest increase of the three cases. Generally speaking, addition of these $CO_2$ capture systems doubles the cost of electricity.

### 6.2 Cost of $CO_2$ Avoidance

Table 6-2 illustrates the cost of $CO_2$ avoidance for the three cases. The cost of $CO_2$ avoided is calculated as follows:

$$\text{Cost of } CO_2 \text{ avoided} = \frac{\left(\text{COE}_{\text{with capture}} - \text{COE}_{\text{without capture}}\right)}{\left(CO_2 \text{ emissions with capture - } CO_2 \text{ emissions without capture}\right)}$$

$$\frac{\$}{\text{tonne } CO_2} [=] \frac{\dfrac{cents}{kWh} \times \dfrac{1000 \text{ kW}}{1 \text{ MW}} \times \dfrac{1\$}{100 \text{ cents}}}{\dfrac{\text{tonne } CO_2}{\text{MWh}}}$$

As shown in the table, the base case cost of $CO_2$ avoidance is 67.2 \$/tonne $CO_2$. The MEA / PZ double matrix cost of $CO_2$ avoidance is 60.2 \$/tonne $CO_2$, which represents a 10% decrease in cost of avoided $CO_2$ emissions. The MDEA / PZ double matrix cost of $CO_2$ avoidance is 55.0 \$/tonne $CO_2$ , which represents a savings of 18% in cost of avoided $CO_2$ emissions. Thus the range of cost savings achieved through use of the advanced solvent formulations and process configurations studied in this effort range from 10 to 18%.

### 6.3 Sensitivity to Plant Size

The DOE Systems Analysis Guidelines suggest that technologies and their costs be evaluated for different base plant sizes ranging from 200 MW to 1000 MW (DOE 2005). The size of these amine units is so large that multiple parallel trains are required due to size limitations for individual pieces of equipment (e.g. absorbers, pumps, heat exchangers, etc.). Therefore, smaller and larger plants would use additional or fewer amine trains and increase or decrease the pieces of equipment in equipment banks. The capital and operating costs scale linearly with respect to gross generating capacity in the range of 200 to 1,000 MW.

### Table 6-2. Cost of $CO_2$ Avoided

| Description | Units | No Capture | MEA Base Case | MEA / PZ Double Matrix | MDEA / PZ Double Matrix |
|---|---|---|---|---|---|
| Plant Net Electrical Capacity - without capture | MWe | 471 | 471 | 471 | 471 |
| Cost of electricity - without capture | c/kWh | 5.0 | 5.0 | 5.0 | 5.0 |
| $CO_2$ emitted - without capture | tonne/h | 462 | 462 | 462 | 462 |
| | tonne/MWh | 0.98 | 0.98 | 0.98 | 0.98 |
| Plant Net Electrical Capacity - with capture | MWe | | 339 | 356 | 378 |
| Cost of electricity - with capture | c/kWh | | 10.7 | 10.1 | 9.7 |
| $CO_2$ emitted - with capture | tonne/h | | 46 | 46 | 46 |
| | tonne/MWh | | 0.14 | 0.13 | 0.12 |
| Cost of $CO_2$ avoided | $/tonne | | 67.20 | 60.19 | 55.05 |
| Percent reduction in cost of $CO_2$ avoided | % | | | 10.4 | 18.1 |

*Note: "=" indicates a value equal to the MEA base case.*

**References (Section 6)**

Department of Energy (DOE) National Energy Technology Laboratory (NETL). "Carbon Capture and Sequestration Systems Analysis Guidelines", April 2005.

## 7.0 SUMMARY AND CONCLUSIONS

This section summarizes the work completed under this Phase I SBIR project and presents the major findings and conclusions.

This project furthers previous work done by Trimeric and the University of Texas at Austin that proposed novel solvent formulations and process configurations. These novel systems could theoretically reduce energy costs of amine-based $CO_2$ capture at coal-fired power plants by 5% to 20% when compared to conventional systems employing monoethanolamine (MEA). The specific technical objectives for Phase I of this SBIR project include the following:

- Establish the two most promising systems of solvent formulation and process scheme,
- Estimate the capital and operating costs of these top two systems,
- Compare these economics with baseline MEA configuration and with Office of Fossil Energy (OFE) targets,
- Resolve how the amine and compression systems will integrate with the power plant, and
- Select the best process configuration and solvent formulation for pilot testing in Phase II.

The Phase I work plan included six tasks to achieve the above technical objectives. First, a process screening study selected the two most promising advanced systems based on the calculated energy requirements for numerous combinations of solvent formulations and process configurations. Second, simulations were prepared for the inlet gas train, amine units, and compression trains in order to prepare heat and material balances. Third, equipment was sized and selected based on the simulation output. Input from industry sponsors was used to make appropriate assumptions and calculations for integrating the $CO_2$ capture system with the main power facility. Then, capital and operating costs were estimated. Finally, cost of electricity and cost of avoided $CO_2$ emissions were calculated in order to compare the two advanced cases with the updated base case.

An updated base case and two advanced cases were evaluated in detail:

| Case Name | Solvent | Configuration |
|---|---|---|
| Base Case | 7 m MEA | Conventional |
| MEA / PZ double matrix | 7 m MEA, 2 m PZ | Double Matrix |
| MDEA / PZ double matrix | Proprietary concentrations | Double Matrix |

*Note: "m" equals molal.*

The design basis for these evaluations was a 500 MW gross conventional coal-fired power plant using Illinois #6 subbituminous coal. A wet flue gas desulfurization (FGD) unit was assumed to be located upstream of the $CO_2$ capture unit. The target $CO_2$ removal is 90%. Any captured $CO_2$ is delivered at pipeline pressure (15.2 MPa, 2200 psia).

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

- Estimates for the reductions in the cost of $CO_2$ capture ($/tonne $CO_2$ avoided) when compared to the base case MEA system ranged from 10 to 18 percent among the cases;

- Estimates for increases in the cost of electricity were 113% for the MEA base case, 102% for the MEA / PZ double matrix, and 95% for the MDEA / PZ double matrix. The base electricity cost used in this study was 5 cents/kWh.

- The configuration with the lowest estimated cost per tonne avoided was the MDEA / PZ double matrix (55.05 $/tonne $CO_2$ avoided);

- The derating due to $CO_2$ capture could be reduced by an estimated 13 to 30% (17 to 39 MWe) by employing advanced solvent formulations and process configurations; and

- Estimated reboiler steam requirements were reduced by 20 to 44 percent, which is desirable from the utility operating perspective though the capital costs to achieve these changes are large.

These results represent improvements in the economics; however, the results do not meet the DOE's goal of achieving $CO_2$ capture with less than a 20 % increase in the cost of electricity. Consequently, this Phase I report marks the end of this particular SBIR research project.