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Details of A Technical, Economic and Environmental Assessment of Amine-based CO\textsubscript{2} Capture Technology for Power Plant Greenhouse Gas Control

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1. INTRODUCTION

1.1. Technology Options for CO₂ Capture

A wide range of technologies currently exist for separation and capture of CO₂ from gas streams, although they have not been designed for power plant scale operations (Desideri and Corbelli, 1998). They are based on different physical and chemical processes including absorption, adsorption, membranes and cryogenics. Figure 1 and Table 1 briefly summarizes the salient features of these technology options (Riemer, et al., 1993; Hendriks, 1994; Mimura et al., 1999; Jeremy, 2000; Audus, 2000). The choice of a suitable technology (which mainly depends on the power plant technology) depends upon the characteristics of the gas stream from which CO₂ needs to be separated. Future power plants may be designed so as to separate out CO₂ from coal before combustion (using coal-gasification systems), or they may employ pure oxygen combustion instead of air so as to obtain a concentrated CO₂ stream for treatment. Figure 2 shows the variety of power plant fuels and technologies that affect the choice of CO₂ capture systems. In this report, post-combustion capture of CO₂ from flue gas streams of conventional power plant using amine-based absorption systems has been considered.

![Figure 1. Technology Options for CO₂ Separation and Capture](image1)

![Figure 2. Technology Options for Fossil-Fuel based Power Generation](image2)
Table 1. Comparison of Technology Options for CO2 Separation and Capture

<table>
<thead>
<tr>
<th>Technology Option</th>
<th>System Requirements</th>
<th>1.2. Advantages</th>
<th>Problems/Drawbacks</th>
</tr>
</thead>
<tbody>
<tr>
<td>Absorption (Chemical)</td>
<td>Absorber and stripper sections Chemical solvent (e.g. MEA, HPC)</td>
<td>Suitable for dilute CO2 streams (typical flue gas from power plants) Operates at ordinary T &amp; P Commercially available, proven technology</td>
<td>The heat of solvent regeneration is very high Significant solvent losses due to acidic impurities in the gas stream</td>
</tr>
<tr>
<td>Absorption (Physical)</td>
<td>Absorber and stripper sections Physical solvent (e.g. Selexol)</td>
<td>Less energy required Solvents are less susceptible to the impurities in the gas stream</td>
<td>Requires high operating pressure Works better with gas streams having high CO2 content</td>
</tr>
<tr>
<td>Adsorption</td>
<td>Adsorber bed(s)</td>
<td>Very high CO2 removal is possible</td>
<td>Requires very high operating pressures Costly</td>
</tr>
<tr>
<td>Membranes</td>
<td>Membrane filter(s)</td>
<td>Upcoming, promising technology Space efficient</td>
<td>Requires very high operating pressures May require multiple units and recycling due to lower product purity Very costly</td>
</tr>
</tbody>
</table>

1.3. Post-combustion amine-based absorption of CO2 from flue gases

Today the 300 GW of coal-fired power generation capacity in the U.S. provides 51% of all power generation and accounts for 79% of carbon emissions coming from electric utilities. Even with the expected growth in natural gas for new generating capacity, coal’s share of the electricity supply is still projected to be about 44% in 2020, and higher in the absolute amount compared to today, according to the latest DOE projections [20]. Natural gas use is projected to account for 31% of power generation in 2020. Thus, any serious policies to reduce CO2 emissions during the next two decades must consider not only the technology options for new power plants (which is the case commonly discussed in the literature), but also the retrofitting of existing coal and natural gas plants which will continue to operate for several decades to come. This medium-term intervention to reduce CO2-emissions has received very little attention to date.

In current systems which use air for combustion, post-combustion CO2 separation from the flue gas stream has to be carried out. Past studies have shown that amine-based CO2 absorption systems are the most suitable for combustion-based power plants for the following reasons
These systems are effective for dilute CO\textsubscript{2} streams, such as coal combustion flue gas which typically contains about 10\%-12 \% CO\textsubscript{2} by volume.

Amine-based CO\textsubscript{2} capture systems are a proven technology that are commercially available today.

Amine-based systems are similar to other end-of-the-pipe environmental control units used at power plants. These units are operated at ordinary temperature and pressure.

A major effort is being made worldwide to improve this process in the light of its potential role in CO\textsubscript{2} abatement. Thus one can anticipate future benefits from technology advances.

1.3. Model Configuration Options

For post-combustion CO\textsubscript{2} capture from flue gas, the amine-based CO\textsubscript{2} capture system, which is the current commercially available technology, has been chosen for this model. There is a major R&D effort going on worldwide to improve this technology – mainly to reduce the high energy penalty of this technology. A substantial part of the energy requirement consists of heat or steam requirement for sorbent regeneration. Depending upon how this steam is supplied, there are three configuration options available. These are shown graphically in Figure 3 and described below.

**Base plant de-rating:** Here, the low-pressure (LP) steam is extracted from the steam cycle of the power plant and supplied to the reboiler for sorbent regeneration. Extraction of steam leads to loss of power generation capacity, and the net plant output decreases substantially. In case of a new plant to be designed with CO\textsubscript{2} capture system, it is possible to optimally design the steam cycle to take care of the steam requirement of the amine system, and proper heat integration may help in reducing the energy penalty. In case of an existing coal plant to be retrofitted with amine system, optimal heat integration may not be achievable, and is likely to lead to much higher energy penalty due to steam extraction.

**Auxiliary Natural Gas Boiler (w/ Steam Turbine):** Another potential option to provide the energy for the amine system is by adding an auxiliary NG-fired boiler. Often it would be combined with a steam turbine which could generate some additional power (mainly used to supply electrical energy demand of CO\textsubscript{2} capture unit), and the LP steam would be then used for sorbent regeneration. Thus, the original steam cycle of the power plant remains unperturbed and the net power generation capacity of the power plant does not get adversely affected. Again, it comes at an additional cost of capital requirement for the boiler (and turbine) and the cost of supplemental NG fuel. Also, the combustion of NG leads to additional CO\textsubscript{2} emissions (and NO\textsubscript{x} emissions). So, there are at least two possible sub-options available

1. The flue gas from the auxiliary boiler is cooled down to acceptable exhaust temperature and then directly vented to the atmosphere. Here, the net CO\textsubscript{2} capture efficiency of the system is substantially lowered because of the additional CO\textsubscript{2} emissions from NG boiler. Also, the total NO\textsubscript{x} emissions may exceed the allowable levels of emission. So, the flue gas from the auxiliary NG boiler may have to be treated for NO\textsubscript{x} removal before venting.

2. The CO\textsubscript{2} capture system maybe designed so as to capture CO\textsubscript{2} from the additional flue gas as well. In this case, the secondary flue gas stream (after cooling and NO\textsubscript{x} removal, if required) maybe merged with main flue gas stream, before it enters the CO\textsubscript{2} capture system. If the NG fuel contains H\textsubscript{2}S, the secondary flue gas may have to be treated for SO\textsubscript{x} removal as well. The basic purpose of the auxiliary NG boiler is to provide the
steam required for sorbent regeneration. With higher amount of flue gas to be treated (and more \( \text{CO}_2 \) to be captured), the amine-system would require more steam and thus a bigger auxiliary NG boiler would be required (which means more secondary flue gas!). An optimal size of auxiliary NG boiler maybe determined by an iterative calculation procedure, so that it matches the sorbent regeneration steam requirement of the \( \text{CO}_2 \) capture system treating the total flue gas. Thus, the \( \text{CO}_2 \) capture level is maintained to the originally desired level, but it often requires substantially big auxiliary NG boiler facility. This may not be always practically feasible (space constraints for retrofit applications, fuel availability, etc.) and economically viable (higher capital cost of building a bigger \( \text{CO}_2 \) capture system as well as an auxiliary boiler, higher O&M costs etc.). In the present version of IECM, this configuration option is not included.

**Figure 3. \( \text{CO}_2 \) Capture Plant Configuration Options**

In terms of the configuration of the \( \text{CO}_2 \) capture system shown in Figure 3, the user can make the following choices as well.
**Direct contact cooler:** The default setting in IECM chooses to include a DCC to cool the flue gas before it enters the amine system. The temperature of the flue gas affects the absorption reaction (absorption of CO$_2$ in MEA sorbent is an exothermic process favored by lower temperatures). Also, the volumetric flow rate of the flue gas stream, which is a key determinant of the sizes of various equipments (direct contact cooler, flue gas blower, absorber), is directly related to the flue gas temperature. Hence lower flue gas temperature is desired. The typically acceptable range of flue gas temperature is about 50-60 °C. If the flue gas is coming from wet sulfur scrubber, additional DCC may not be required. But in case of flue gas from NG-fired boiler, which often does not pass through a sulfur scrubber, DCC is a must.

**Choice of sorbent:** At this time, MEA is the default sorbent used in the system and the nominal values of various parameters are based on a process simulation model that uses MEA. As always, the users can overwrite the nominal values of these parameters if they wish to use a different sorbent (and have the relevant data). In future, the model can adopt a different sorbent by providing the appropriate values for the key parameters.

**CO$_2$ transportation:** The default mode of CO$_2$ transportation is via pipelines. The user can specify the distance over which CO$_2$ needs to be carried to, and the unit cost of CO$_2$ transportation. This module maybe expanded in future to include detailed parameters about pipeline transport and also other transport options.

**CO$_2$ storage/ disposal:** The default option for CO$_2$ disposal is underground geological storage. A nominal cost of $5/ tonne CO$_2$ has been suggested, which can be changed the user to match the specific details about the location. If CO$_2$ is being used as a byproduct for EOR or ECBM activity, it may generate some revenue. This module, which is represented by a single cost parameter, maybe expanded in future to include details about the various storage/ disposal options.

### 2. AMINE-BASED CO$_2$ CAPTURE SYSTEMS

#### 2.1. Historical Developments

Combustion of fossil fuels in air leads to a gaseous product stream that mainly contains nitrogen, carbon dioxide, water vapor and small quantities of many other gases. Depending upon the carbon content of the fuel (and the quantity of air used for combustion of the fuel), the flue gas stream may contain as high as 15% CO$_2$ and is an obvious source of CO$_2$ available at no cost. The whole idea of separating CO$_2$ from flue gas streams started in 1970’s, not with concern about the greenhouse effect, but as a possibly economic source of CO$_2$, mainly for enhanced oil recovery (EOR) operations. Even today, about 80% of CO$_2$ production is used for EOR (Chapel et al., 1999). Several commercial CO$_2$ capture plants were constructed in the US in the late 1970’s and early 1980’s (Kaplan, 1982; Pauley, et al., 1984). CO$_2$ was also produced for other industrial applications such as carbonation of brine and production of products like dry ice, urea and beverages. Some of these plants are still in operation today. But all these plants are much smaller (in terms of tonnage of CO$_2$ handled) than a typical power plant. Figure 4 gives a rough idea about the various industrial applications of CO$_2$ capture technologies and their relative magnitude of operations. The first commercial CO$_2$ sequestration facility started in Norway in September 1996 in response to a Norwegian carbon tax. Since then, Statoil has been storing about 1 million tonnes of CO$_2$ per year from the Sleipner West gas field into a sandstone aquifer 1000 m beneath the North Sea (USDOE, 1999; Statoil, 2001). The international research community is closely monitoring this facility.
All these plants capture CO\textsubscript{2} with processes based on chemical absorption using a monoethanolamine (MEA) based sorbent. MEA is an organic chemical belonging to the family of compounds known as amines. It was developed over 60 years ago as a general, non-selective sorbent to remove acidic gas impurities (e.g. H\textsubscript{2}S, CO\textsubscript{2}) from natural gas streams (Herzog, 1999). The process was then adapted to treat flue gas streams for CO\textsubscript{2} capture. Dow Chemical Co. (and later Fluor Daniel Inc.), Kerr-McGee Chemical Corp. and ABB Lummus Crest Inc., were some of the initial developers of MEA-based technology for CO\textsubscript{2} capture. About 75%-95% CO\textsubscript{2} may be captured using this technology to yield a fairly pure (>99%) CO\textsubscript{2} product stream.

Today there are two main MEA-based processes available for commercial CO\textsubscript{2} recovery plants: the Fluor Daniel Econamine FG process and the ABB Lummus Crest MEA process (Wong et al., 2000). Data for the Econamine FG process are more readily available. So, the performance and cost model is based on this process, which uses 30% w/w MEA solution with an oxygen inhibitor. The inhibitor helps in two ways – reduced sorbent degradation and reduced equipment corrosion (Chapel et al., 1999). It may be noted that this process is not applicable to reducing gas streams that contain large amounts of CO and H\textsubscript{2}, or contain more than 1 ppm of H\textsubscript{2}S, or contain less than 1% O\textsubscript{2} v/v. On the other hand, the ABB Lummus Crest process uses a 15%-20% w/w MEA solution without any inhibitor (Marion et al., 2001). This technology can capture more than 96% of the CO\textsubscript{2} from flue gases, but the lower sorbent concentration leads to economic disadvantages in terms of greater capital requirements (due to larger equipment size) and higher energy requirements (due to higher amount of dilution water per unit of sorbent).
2.2. Process Description

A continuous scrubbing system is used to separate CO₂ from a gaseous stream. The system consists of two main elements, an absorber, where CO₂ is absorbed into a sorbent and a regenerator (or stripper), where CO₂ is released (in concentrated form) and the original sorbent is recovered. Chemical absorption systems tend to be more efficient than the other systems shown in Appendix A, as the process is accompanied by a chemical reaction that enhances the overall mass transfer from gas phase to liquid phase.

In a power plant application (Figure 5) cooled flue gases flow vertically upwards through the absorber countercurrent to the absorbent (MEA in a water solution, with some additives). The MEA reacts chemically with the CO₂ in the flue gases to form a weakly bonded compound (carbamate). The scrubbed gases are then washed and vented to the atmosphere. The CO₂-rich solution leaves the absorber and passes through a heat exchanger, then further heated in a reboiler using low-pressure steam. The weakly bonded compound formed during absorption is broken down by the application of heat, regenerating the sorbent, and producing a concentrated CO₂ stream. The hot CO₂-lean sorbent is then returned to the heat exchanger, where it is cooled, then sent back to the absorber. Some fresh MEA is added make up for losses incurred in the process.

![Flow Sheet for CO₂ Capture from Flue Gases using Amine-based System](image)

*Figure 5. Flow Sheet for CO₂ Capture from Flue Gases using Amine-based System*

The CO₂ product is separated from the sorbent in a flash separator, and then taken to the drying and compression unit. It is compressed to very high pressures (about 2000 psig) so that it is liquefied and easily transported to long distances to the designated storage or disposal facility.
2.3. Process Chemistry

The process chemistry is complex, but the main reactions taking place are [26]

\[ \text{CO}_2 \text{ Absorption: } 2 \text{R-NH}_2 + \text{CO}_2 \rightarrow \text{R-NH}_3^+ + \text{R-NH-COO} \]

\[ \text{MEA Regeneration: } \text{R-NH-COO}^- + \text{R-NH}_3^+ + (\text{Heat}) \rightarrow \text{CO}_2 + 2 \text{R-NH}_2 \]

Pure MEA (with R = HO-CH₂CH₂) is an “unhindered” amine that forms a weakly bonded intermediate called “carbamate” that is fairly stable. Only half a mole of CO₂ is absorbed per mole of amine, as shown in the CO₂ absorption equation above. On application of heat, this carbamate dissociates to give back CO₂ and amine sorbent, as shown in the second equation above. Since the carbamate formed during absorption is quite stable, it takes lot of heat energy to break the bonds and to regenerate the sorbent.

For other “hindered” amines (e.g., where R is a bulky group), the carbamate formed is not stable, and an alternate reaction leads to formation of bicarbonate ions and hence a higher theoretical capacity of one mole of CO₂ per mole of amine, as shown in the CO₂ absorption equation below (Hezorg et al., 1997; Sartori, 1994).

\[ \text{CO}_2 \text{ Absorption: } \text{R-NH}_2 + \text{CO}_2 + \text{H}_2\text{O} \rightarrow \text{R-NH}_3^+ + \text{HCO}_3^- \]

\[ \text{MEA Regeneration: } \text{HCO}_3^- + \text{R-NH}_3^+ + (\text{less Heat}) \rightarrow \text{CO}_2 + \text{H}_2\text{O} + \text{R-NH}_2 \]

The regeneration of these amines requires lesser amount of heat energy as compared to the unhindered amines. But the CO₂ uptake rate of hindered amines is very low. Efforts are underway to formulate better sorbents by combining favorable properties of these two groups of amines.

2.4. Process Equipment

The CO₂ capture and separation system consists of the following capital equipment

**Direct contact cooler**: The flue gases coming out of a power plant are quite hot. The temperature of flue gas may be ranging from as low as 60 deg. C (in case of coal-fired power plants with wet SO₂ scrubbers) to more than 550 deg. C (in case of natural gas fired simple cycle power plants). It is desirable to cool down the flue gases to about 45-50 deg. C, in order to improve absorption of CO₂ into the amine sorbent (the absorption being an exothermic process is favored by low temperatures), to minimize sorbent losses (higher temperature may lead to sorbent losses due to evaporation and degradation), and to avoid excessive loss of moisture with the exhaust gases. In case of gas-fired power plants or majority of coal-fired power plants that do not have wet scrubbers for SO₂ removal, a direct contact cooler has to be installed to bring down the temperature of the flue gas stream to acceptable levels. In case of coal-fired power plant applications that have a wet FGD (flue gas desulfurization) unit upstream of the amine system, the wet scrubber helps in substantial cooling of the flue gases, and additional cooler may not be required.

**Flue gas blower**: The flue gas has to overcome a substantial pressure drop as it passes through a very tall absorber column, countercurrent to the sorbent flow. Hence the cooled flue gas has to be pressurized using a blower before it enters the absorber.

**Absorber**: This is the vessel where the flue gas is made to contact with the MEA-based sorbent, and some of the CO₂ from the flue gas gets dissolved in the sorbent. The column may be plate-
type or a packed one. Most of the CO\(_2\) absorbers are packed columns using some kind of polymer-based packing to provide large interfacial area.

**Rich/lean cross heat exchanger:** The CO\(_2\)-loaded sorbent needs to be heated in order to strip off CO\(_2\) and regenerate the sorbent. On the other hand, the regenerated (lean) sorbent coming out of the regenerator has to be cooled down before it could be circulated back to the absorber column. Hence these two sorbent streams are passed through a cross heat exchanger, where the rich (CO\(_2\)-loaded) sorbent gets heated and the lean (regenerated) sorbent gets cooled.

**Regenerator:** This is the column where the weak intermediate compound (carbamate) formed between the MEA-based sorbent and dissolved CO\(_2\) is broken down with the application of heat and CO\(_2\) gets separated from the sorbent to leave reusable sorbent behind. In case of unhindered amines like MEA, the carbamate formed is stable and it takes large amount of energy to dissociate. It also consists of a flash separator where CO\(_2\) is separated from most of the moisture and evaporated sorbent, to give a fairly rich CO\(_2\) stream.

**Reboiler:** The regenerator is connected with a reboiler which is basically a heat exchanger where low-pressure steam extracted from the power plant is used to heat the loaded sorbent.

**Steam extractor:** In case of coal-fired power plants that generate electricity in a steam turbine, a part of the LP/IP steam has to be diverted to the reboiler for sorbent regeneration. Steam extractors are installed to take out steam from the steam turbines.

**MEA reclaimer:** Presence of acid gas impurities (SO\(_2\), SO\(_3\), NO\(_2\) and HCl) in the flue gas leads to formation of heat stable salts in the sorbent stream, which can not be dissociated even on application of heat. In order to avoid accumulation of these salts in the sorbent stream and to recover some of this lost MEA sorbent, a part of the sorbent stream is periodically distilled in this vessel. Addition of caustic helps in freeing of some of the MEA. The recovered MEA is taken back to the sorbent stream while the bottom sludge (reclaimer waste) is sent for proper disposal.

**Sorbent processing area:** The regenerated sorbent has to be further cooled down even after passing through the rich/lean cross heat exchanger using a cooler, so that the sorbent temperature is brought back to acceptable level (about 40 deg C). Also, in order to make up for the sorbent losses, a small quantity of fresh MEA sorbent has to be added to the sorbent stream. So, the sorbent processing area primarily consists of sorbent cooler, MEA storage tank, and a mixer. It also consists of an activated carbon bed filter that adsorbs impurities (degradation products of MEA) from the sorbent stream.

**CO\(_2\) drying and compression unit:** The CO\(_2\) product may have to be carried to very long distances via pipelines. Hence it is desirable that it does not contain any moisture in order to avoid corrosion in the pipelines. Also, it has to be compressed to very high pressures so that it gets liquefied and can overcome the pressure losses during the pipeline transport. The multi-stage compression unit with inter-stage cooling and drying yields a final CO\(_2\) product at the specified pressure (about 2000 psig) that contains moisture and other impurities (e.g. N\(_2\)) at acceptable levels.

**CO\(_2\) transport facility:** The CO\(_2\) captured at the power plant site has to be carried to the appropriate storage/disposal site. Considering the scale of the operation (thousands of tonnes of CO\(_2\) per day), pipelines seems to be the obvious mode of transportation. In case of retrofit applications, where construction of new pipelines might be prohibitively expensive (and
questionable in terms of public acceptance, especially in densely populated regions), transport via tankers may be considered. There is fair amount of industrial experience and expertise in the field of the construction (and operation) of pipelines for CO\textsubscript{2} transport. Recently, a 325-km pipeline carrying CO\textsubscript{2} from the Great Plains Synfuels Plant in Beulah, North Dakota (owned by Dakota Gasification Company of Bismarck, North Dakota) to the Weyburn oil fields in Saskatchewan, Canada went operational (PanCanadian, 2001).

**CO\textsubscript{2} disposal facility:** Once the CO\textsubscript{2} is captured, it needs to be securely stored (sequestered). Again, there are a wide range of options potentially available (see Fig. A-3 and Table A-2 in Appendix A). Geologic formations such as underground deep saline reservoirs, depleted oil and gas wells, and abandoned coal seams are some of the potentially attractive disposal sites [14-16]. Ocean disposal and terrestrial sinks are additional options being studied [17-18]. The distance to a secure storage site and the availability and cost of transportation infrastructure also affect the choice of disposal option. In general, studies indicate that geologic formations are the most plentiful and attractive option for U.S. power plants [19]. Transport of CO\textsubscript{2} to a storage site is typically assumed to be via pipeline. While the economic costs of CO\textsubscript{2} storage appear to be low, the social and political acceptability of different options are not yet clear.

### 2.5. Limitations of the MEA Process

Although MEA-based absorption process is the most suitable technology available for capture of CO\textsubscript{2} from power plant flue gases, it has its own drawbacks. The main problems maybe enlisted as follows

**Energy Penalty:** The stable carbamate ion requires substantial energy to break the bonds. So, a large amount of heat is required to regenerate the sorbent. Substantial energy also is needed to compress the captured CO\textsubscript{2} for pipeline transport to a storage site. This heat and electricity requirement reduces the net efficiency of the power plant if it is extracted internally (by de-rating the power plant). Alternatively, a much bigger power plant needs to be built in order to achieve the same “net” power generation capacity, as it would have been without CO\textsubscript{2} capture.

**Loss of Sorbent:** Some of the sorbent is lost during the process because of a variety of reasons including mechanical, entrainment, vaporization and degradation (Stewart and Lanning, 1994). All the sorbent entering the stripper does not get regenerated. Flue gas impurities, especially oxygen, sulfur oxides and nitrogen dioxide react with MEA to form heat-stable salts, thus reducing the CO\textsubscript{2}-absorption capacity of the sorbent. Proprietary inhibitors are available that make the sorbent tolerant to oxygen. Flue gas NO\textsubscript{x} is not a major problem since nitric oxide (NO) is the predominant form (~ 90-95%) of total NO\textsubscript{x} in the flue gas, and does not react with inhibited amines (Suda et al., 1992; Leci, 1996). But, SO\textsubscript{2} does degenerate MEA sorbent, so very low inlet concentrations (10 ppm) are desirable to avoid excessive loss of sorbent. However, untreated flue gases of coal-fired power plants contain about 700 to 2500 ppm SO\textsubscript{2} (plus roughly 10-40 ppm NO\textsubscript{2}). The interaction of SO\textsubscript{2} with CO\textsubscript{2} control system is thus particularly important. The heat-stable salts that are formed may be treated in a side stream MEA-reclaimer, which can regenerate some of the MEA. Technologies such as electrodialysis are also being proposed for this purpose (Yagi et al., 1992).

**Corrosion:** Corrosion control is very important in amine systems processing oxygen-containing gases. In order to reduce corrosion rates, corrosion inhibitors, lower concentrations of MEA, appropriate materials of construction and mild operating conditions are required (Barchas and Davis, 1992).
3. PERFORMANCE MODEL DEVELOPMENT

A number of previous studies have reported some cost and/or performance data for specific amine-based systems, including hypothetical applications to coal-fired power plants (Smelster, 1991; Kohl and Nielsen, 1997; Chapel et al., 1999; Herzog, 1999; Chakma and Tontiwachwuthikul, 1999; Simbeck, 1999; Marion et al., 2001). However, there are no generally available process models that can be used or modified for detailed studies of CO₂ removal options. Cost data also are relatively limited and often incomplete.

The rate of removal of CO₂ from flue gas using an amine scrubber depends on the gas-liquid mass transfer process. The chemical reactions that permit diffusion of CO₂ in the liquid film at the gas-liquid interface enhance the overall rate of mass transfer. So, the CO₂ removal efficiency in the absorber is a function of various parameters that affect the gas-liquid equilibrium (e.g., flow rates, temperature, pressure, flue gas composition, MEA concentration, equipment design, etc.). Absorption of CO₂ in an alkaline medium may be considered as a first order reaction. Higher CO₂ concentration thus improves the efficiency of the absorption system. Even at low concentration of CO₂, MEA has great affinity for CO₂. The solubility of CO₂ in MEA is much higher as compared to many other conventional solvents. Similarly, the design of and conditions in the regenerator affect the energy requirement and the overall performance of the system.

3.1. Process Simulation Tool

Two process simulators, viz. ASPEN-Plus and ProTreat have been used to derive the performance equations. The CO₂ capture and separation model is based on the ProTreat simulations while the CO₂ compression model is based on the ASPEN-Plus simulations.

- **ProTreat** is a software package for simulating processes for the removal of H₂S, CO₂, and mercaptans from a variety of high and low pressure gas streams by absorption into thermally regenerable aqueous solutions containing one or more amines. The ProTreat package makes exclusive use of a column model that treats the separation as a mass transfer rate process.

- **ASPEN-Plus** is a powerful process engineering tool for the design and steady-state simulation and optimization of process plants.

3.2. Methodology

A large number of process simulation runs have been conducted to cover a reasonable range of values for the key parameters. The details are presented in the following sections.

3.2.1. ProTreat Simulation Runs for CO₂ capture and separation from flue gas

The CO₂ capture and separation system consists of a flue gas compressor, cooler, absorber, heat exchangers, regenerator, sorbent circulation pumps etc. Many parameters were varied in the ProTreat model. Table 2 summarizes the parameters that were changed and the ranges for each parameter.

<table>
<thead>
<tr>
<th>No.</th>
<th>Parameter</th>
<th>Type</th>
<th>Units</th>
<th>Range</th>
</tr>
</thead>
<tbody>
<tr>
<td>1</td>
<td>CO₂ content in flue gas (y_{CO2})</td>
<td>Input</td>
<td>mole %</td>
<td>3.5-13.5</td>
</tr>
<tr>
<td></td>
<td>Parameter Description</td>
<td>Type</td>
<td>Unit</td>
<td>Range</td>
</tr>
<tr>
<td>---</td>
<td>------------------------------------------</td>
<td>--------</td>
<td>-------------</td>
<td>-------------</td>
</tr>
<tr>
<td>2</td>
<td>Flue gas flow rate (G)</td>
<td>Input</td>
<td>kmole/hr</td>
<td>9000-24000</td>
</tr>
<tr>
<td>3</td>
<td>Inlet flue gas temperature (T_{fg})</td>
<td>Input</td>
<td>deg C</td>
<td>40-65</td>
</tr>
<tr>
<td>4</td>
<td>MEA concentration (C)</td>
<td>Input</td>
<td>wt %</td>
<td>15–40</td>
</tr>
<tr>
<td>5</td>
<td>Sorbent flow rate (L)</td>
<td>Input</td>
<td>kmole/hr</td>
<td>16000-70000</td>
</tr>
<tr>
<td>6</td>
<td>L/G</td>
<td>Input</td>
<td>-</td>
<td>0.73-5.56</td>
</tr>
<tr>
<td>7</td>
<td>Reboiler heat duty (Q)</td>
<td>Input</td>
<td>GJ/hr</td>
<td>95-664</td>
</tr>
<tr>
<td>8</td>
<td>Q/L</td>
<td>Input</td>
<td>MJ/kmole</td>
<td>2.4-22.5</td>
</tr>
<tr>
<td>9</td>
<td>CO₂ capture efficiency</td>
<td>Output</td>
<td>%</td>
<td>41.2-99.9</td>
</tr>
<tr>
<td>10</td>
<td>CO₂ product flow rate</td>
<td>Output</td>
<td>kmole/hr</td>
<td>333-2840</td>
</tr>
<tr>
<td>11</td>
<td>Lean sorbent CO₂ loading</td>
<td>Output</td>
<td>moleCO₂/mole MEA</td>
<td>0.05-0.34</td>
</tr>
<tr>
<td>12</td>
<td>Rich sorbent CO₂ loading</td>
<td>Output</td>
<td>moleCO₂/mole MEA</td>
<td>0.27-0.55</td>
</tr>
<tr>
<td>13</td>
<td>Absorber diameter</td>
<td>Output</td>
<td>ft</td>
<td>26-42</td>
</tr>
<tr>
<td>14</td>
<td>Regenerator diameter</td>
<td>Output</td>
<td>ft</td>
<td>12-42</td>
</tr>
<tr>
<td>15</td>
<td>Exhaust flue gas temperature</td>
<td>Output</td>
<td>deg C</td>
<td>40.4-71.6</td>
</tr>
</tbody>
</table>

The following set of parameters related to the design/configuration of the CO₂ capture system were held constant:

- Absorber height: 40 ft
- Absorber packing: Rasching rings, metallic, 1-inch packing size
- Inlet flue gas pressure: 3 psi
- Solvent pumping pressure: 30 psi
- Number of trays in regenerator: 24 (tray spacing = 2 ft, weir height = 3 inches)

### 3.2.2. ASPEN-Plus Simulation Runs for CO₂ Compression

The concentrated CO₂ product stream obtained from sorbent regeneration is compressed and dried using a multi-stage compressor with inter-stage cooling. The ASPEN-Plus module used for this simulation consists of 4 stages of compression with inter-stage cooling that deliver the compressed product at 35°C. The compressor efficiency, CO₂ product pressure and purity were used as the main control variables. These parameters were varied over the following ranges:

- Compressor efficiency: 60-100 %
- CO₂ product pressure: 500-2500 psi
- CO₂ stream purity: 99-100%

3.2.3. **Regressions using SAS to derive performance equations**

The IECM uses response-surface models to characterize the performance of various technologies. Simple algebraic equations are derived from the process simulation runs and used as performance equations rather than having a detailed process simulation module inside IECM. The key performance output variables were regressed against all the input variables to obtain linear/logarithmic relationship among them. The data collected from the process simulation runs was used to carry out these multivariate linear regressions using a statistical package called SAS. Only those variables with significance value greater than 0.9995 were retained in the performance equations.

3.3. **Performance Parameters**

A preliminary model was developed to simulate the performance of a CO₂ capture system based on amine (MEA) scrubbing. This CO₂ module was then added to an existing coal-based power plant simulation model (called IECM), described later in this section. Basically, there are three types of input parameters to the CO₂ performance model:

- **Parameters from the “reference plant”:** These include the flow rate, temperature, pressure and composition of the flue gas inlet to the CO₂ absorber, and the gross power generation capacity of the power plant.

- **Parameters to configure the CO₂ system:** The CO₂ module provides a menu of options from which the user may select a CO₂ capture technology, CO₂ product pressure, mode and distance of CO₂ product transport, and CO₂ storage/disposal method. At this stage, a model of the MEA-based absorption system with pipeline transport and geologic sequestration has been developed; other options shown in Appendix A are still under construction.

- **Parameters controlling the performance of the CO₂ system:** The main parameters include the CO₂ capture efficiency, MEA concentration, maximum and lean CO₂ loadings of the solvent, regeneration heat requirement, pressure drop across the system, MEA make-up requirement, pump efficiency, compressor efficiency and several others.

These parameters are used to calculate the solvent flow rate, MEA requirement, and energy penalty of the CO₂ system.

Functional relationships and default values for all model parameters were developed based on engineering fundamentals, a detailed review of the literature, and several contacts with experts in the field. All of these performance parameters directly affect the cost of the system.

Here is a brief description of the various input parameters to the CO₂ system.

3.3.1. **Parameters obtained from the “reference base plant”**

The amine-based CO₂ capture system gets the following inputs from the (reference) base plant:

\[
\text{Gross plant size} = MW_g \\
\text{Net plant size (after env\'l. controls)} = MW_{nctl}
\]
Flue gas composition and flow rate (as entering into the amine system)
This is an array of molar flow rates of different gas components that include N2, O2, H2O, CO2, CO, HCl, SO2, SO3, NO, NO2 and mass flow rate of particulates. The total molar flow rate of the flue gas is G, and the molar fraction of CO2 in the flue gas is yCO2.

- Temperature of flue gas = Tfg
- Plant capacity factor = PCF (%)
- Annual hours of operation = HPY = (PCF/100)*365*24 hrs/yr

3.3.2. Parameters to configure the CO2 system
These are the choices the user can make in order to configure the CO2 capture system.

- Flue gas cooler: Whether to include DCC (default) or excluded
- Sorbent regeneration steam supply: Steam extraction from the base plant (default, internal derating) or Steam generated from an auxiliary NG boiler (w/ ST)
- Mode of CO2 product transportation: Via pipelines (default) or any other means.
- Mode of CO2 storage/disposal: Underground geologic reservoir (default) or EOR or ECBM or Depleted oil/gas wells or Ocean

3.3.3. Parameters controlling the performance of the CO2 system
Parameters controlling the performance of the CO2 system: The numerical values to the input parameters are specified by the user. The intermediate and final output parameters are then derived using the performance equations. It may be noted that the user can override any of these values, but may want to change values of all the relevant parameters to avoid inconsistencies.

**CO2 capture efficiency (ηCO2)**
The overall CO2 capture efficiency of the system is the fraction of CO2 present in the incoming flue gas stream captured in this system.

\[
\eta_{CO2} = \frac{\text{Moles CO2 in} - \text{Moles CO2 out}}{\text{Moles CO2 in}}
\]

Most of studies report the CO2 capture efficiency of the amine-based systems to be 90%, with few others reporting as high as 96% capture efficiency. Here, it has been assumed to be 90% as nominal value, but the user can specify the desired level of CO2 capture efficiency.

**MEA concentration (CMEA)**
The solvent used for CO2 absorption is a mixture of monoethanolamine (MEA) with water. MEA is a highly corrosive liquid, especially in the presence of oxygen and carbon dioxide, and hence needs to be diluted. Today the commercially available MEA-based technology supplied by Fluor Daniel uses 30% w/w MEA solvent with the help of some corrosion inhibitors. Other suppliers, who do not use this inhibitor, prefer to use lower MEA concentrations in the range of 15%-20% w/w. Here we use 30% as the nominal value for the solvent concentration and the user may choose any value between 15-40%.

**Lean solvent CO2 loading (ϕ_min)**
Ideally, the solvent will be completely regenerated on application of heat in the regenerator section. Actually, even on applying heat, not all the MEA molecules are freed from CO2. So, the
regenerated (or lean) solvent contains some “left-over” CO₂. The level of lean solvent CO₂ loading mainly depends upon the initial CO₂ loading in the solvent and the amount of regeneration heat supplied, or alternatively, the regeneration heat requirement depends on the allowable level of lean sorbent loading. Here we use a nominal value of 0.2 based on the values reported in the literature, and the user may specify any desired value in the range (0.05-0.3).

**Liquid to gas ratio (L/G)**
The liquid to gas ratio is the ratio of total molar flow rate of the liquid (MEA sorbent plus water) to the total molar flow rate of flue gas being treated in the absorber. This is one of the parameters derived by the process simulation model.

**Liquid flow rate (L)**
The liquid flow rate is the total molar flow rate of sorbent plus dilution water being circulated in the CO₂ capture system. It is obtained by multiplying (L/G) which is derived from the process simulation model, by the total flue gas flow rate (G) entering the CO₂ capture system.

\[
L = (L/G) \times (G)
\]

**Removal efficiency (η_{acid gas}) and stoichiometric MEA loss (n_{MEA,acidgas})**
As discussed before, MEA is an alkaline solvent that has strong affinity for various acid gases. In fact, gases such as hydrogen chloride and oxides of sulfur are much more reactive towards MEA than carbon dioxide itself. Also, these gases form heat stable salts (HSS) with MEA that can not be regenerated even after application of heat. So, they cause a (permanent) loss of MEA solvent that may be estimated according the stoichiometry of their reactions with MEA. The typical removal efficiencies of these gases in the absorber using MEA solvent designed for 90% removal of CO₂ are given in Table 3.

<table>
<thead>
<tr>
<th>Acid gas</th>
<th>removal efficiency (%)</th>
<th>MEA loss (mole MEA/mole acid gas)</th>
</tr>
</thead>
<tbody>
<tr>
<td>SO₂</td>
<td>η_{SO₂} = 99.5%</td>
<td>n_{MEA, SO₂} = 2</td>
</tr>
<tr>
<td>SO₃</td>
<td>η_{SO₃} = 99.5%</td>
<td>n_{MEA, SO₃} = 2</td>
</tr>
<tr>
<td>NO₂</td>
<td>η_{NO₂} = 25%</td>
<td>n_{MEA, NO₂} = 2</td>
</tr>
<tr>
<td>NO</td>
<td>η_{NO} = 0</td>
<td>n_{MEA, NO} = 0</td>
</tr>
<tr>
<td>HCl</td>
<td>η_{HCl} = 95%</td>
<td>n_{MEA, HCl} = 1</td>
</tr>
</tbody>
</table>

**Temperature of the flue gas entering the CO₂ capture system (T_{fg,in})**
The desirable temperature of the flue gas entering the CO₂ capture system is about 45-50 deg C. If a direct contact cooler is installed upstream of CO₂ capture system, then this temperature level may be achieved. Else, this is same as that obtained from the base plant.

The temperature of the flue gas affects the absorption reaction (absorption of CO₂ in MEA solvent is an exothermic process favored by lower temperatures). Also, the volumetric flow rate
of the flue gas stream, which is a key determinant of the sizes of various equipments (direct contact cooler, flue gas blower, absorber), is directly related to the flue gas temperature.

**Nominal MEA loss (\( \dot{n}_{\text{MEA, \text{nom}}} \))**

MEA is a reactive solvent. In spite of dilution with water and use of inhibitors, a small quantity of MEA is lost through various unwanted reactions, mainly the polymerization reaction (to form long-chained compounds) and the oxidation reaction forming organic acids and liberating ammonia. In general, this nominal loss of MEA is estimated as about 1.5 kgMEA/mton CO\(_2\).

It is also assumed that 50% of this MEA loss is due to polymerization:

\[
\dot{n}_{\text{MEA, polym}} = 50\% \text{ of } \dot{n}_{\text{MEA, nom}}
\]

and the remaining 50% of the MEA loss is due to oxidation to acids:

\[
\dot{n}_{\text{MEA, oxid}} = 50\% \text{ of } \dot{n}_{\text{MEA, nom}}.
\]

**NH\(_3\) Generation (\( n_{\text{NH}_3} \))**

The oxidation of MEA to organic acids (oxalic, formic, etc.) also leads to formation of NH\(_3\). Each mole of MEA lost in oxidation, liberates a mole of ammonia (NH\(_3\)).

\[
\text{Rate of ammonia generation, } n_{\text{NH}_3} = 1 \frac{\text{mole } \text{NH}_3}{\text{mole MEA oxidized}}
\]

**Heat-Stable Salts (HSS)**

The organic acids (product of MEA oxidation) combine with MEA to form some other heat stable salts (HSS). The exact nature of these salts is not known. The most conservative estimate, assuming that the organic acids are mono-basic, is that each mole of organic acid takes up one mole of fresh MEA. [Each mole of MEA lost in oxidation takes up additional mole of MEA in HSS formation.]

\[
\dot{n}_{\text{MEA, organics}} = \left( \frac{1}{1 \text{mole MEA}} \right) \text{mole org. acids}
\]

**Caustic Consumption in Reclaimer (\( \dot{m}_{\text{NaOH}} \))**

Caustic (in the form of NaOH) is added in the reclaimer so that some of the MEA could be regenerated from HSS. \( \dot{m}_{\text{NaOH}} \) is the quantity (mass) of caustic (as NaOH) consumed in MEA reclaimer per tonne of CO\(_2\) captured. A typical value is about 0.13 kg NaOH/mton CO\(_2\).

**Reclaimed MEA**

Caustic regenerates stoichiometric amount of MEA from the HSS in the reclaimer. Each mole of NaOH regenerates 1 mole of MEA, and adds the corresponding Na salt of organic acid to the reclaimer bottoms.

\[
\dot{n}_{\text{MEA, reclaimed}} = \text{no. of moles of MEA reclaimed using caustic}
\]

\[
= \text{no. of moles of caustic added}
\]
\[ \dot{m}_{NaOH} = \dot{n}_{NaOH} \]
\[ \dot{m}_{NaOH} = \dot{n}_{NaOH} \cdot \text{(Molecular Weight of NaOH)} \]
\[ \dot{m}_{NaOH} = \dot{n}_{NaOH} / 40 \]  
(7)

**Removal efficiency for particulates** (\(\eta_{\text{partic}}\))

Amine-based absorption system for CO\(_2\) removal is a wet scrubbing operation. So, it also leads to removal of particulate matter from the flue gas to certain extent. Based on the experience of other scrubbing systems, the removal efficiency for particulates has been assumed to be 50% (which may be a function of particle size distribution).

**Density of sorbent** (\(\rho_{\text{sorbent}}\))

MEA has a density (1.022 g/cc) that is similar to that of water. So, the overall density of the MEA based solvent (with almost 70% water) is assumed to be same as that of water ~1 mton/m\(^3\).

**Activated Carbon** (\(\dot{m}_{\text{act-C}}\))

Activated carbon bed in the solvent circuit helps in removal of long chained/ cyclic polymeric compounds formed from the degenerated MEA. Over a period of time (~3-6 months) the C-bed needs to be replaced (the used bed is sent back to the suppliers). \(\dot{m}_{\text{act-C}}\) is the average amount of activated carbon consumed per tonne of CO\(_2\) captured. Typically, this consumption is estimated to be about 0.075 kg C/ tonne CO\(_2\).

**Total moles of CO\(_2\) captured** (\(n_{CO2}\))

This is the molar flow rate of CO\(_2\) captured from the flue gas. It is obtained by multiplying the total CO\(_2\) content in the inlet flue gas (kmole CO\(_2\) / hr) by the CO\(_2\) capture efficiency of the system.

\[ n_{CO2} = (\eta_{CO2} / 100) \cdot (\text{Moles CO}_2 \text{ in}) = (\eta_{CO2} / 100) \cdot (G \cdot Y_{CO2}) \]

Since the molecular weight of CO\(_2\) is 44, the total amount of CO\(_2\) captured (\(m_{CO2}\), tonne/ hr) is

\[ m_{CO2} = n_{CO2} \cdot (44/1000) \]

**CO\(_2\) product purity** (\(\alpha\))

The final CO\(_2\) product has to meet certain specifications depending upon the mode of transport and final destination. Impurities such as nitrogen are undesirable as they may pose problems during compression and liquefaction of CO\(_2\). In order to avoid corrosion in the pipelines during transport, the moisture levels have to be controlled. The acceptable level of purity of CO\(_2\) product for most of the applications is about 99.8%.

**Reboiler duty per mole of liquid** (\(Q/L\))

This is the total amount of heat energy input required for the regeneration of the sorbent per unit of liquid circulated. This is mainly dependent on lean sorbent loading, CO\(_2\) capture efficiency, MEA concentration and CO\(_2\) content of the flue gas and is derived form the process simulation model.
**Total heat requirement for sorbent regeneration (Q)**
This is the total amount of heat energy required in the reboiler for sorbent regeneration. It is obtained by multiplying \((Q/L)\) which is derived from the process simulation model, by the total sorbent circulation molar flow rate (MEA sorbent plus dilution water) in the CO\(_2\) capture system.

\[
Q = (Q/L) \times (L)
\]

**Unit heat of sorbent regeneration \((q_{\text{regen}})\)**
This is the amount of heat required for the regeneration of the MEA solvent (loaded with CO\(_2\)) in the stripper/ regenerator section. It is expressed as amount of heat (in kJ or Btu) per unit mass (kg or lb) of CO\(_2\) captured. Theoretically, the heat of reaction that needs to be supplied in order to reverse the absorption reaction between CO\(_2\) and MEA is about 825 Btu/ lb CO\(_2\) (i.e. about 1900 kJ/ kg CO\(_2\)). The actual amount of heat required for regeneration of the solvent is much higher (about 2-3 times higher than this theoretical minimum), mainly because of the large amount of latent heat taken up by the dilution water in the solvent. A wide range of numbers have been reported for the regeneration heat requirement of MEA system. Majority of the sources report a heat requirement of about 3800-4000 kJ/kg CO\(_2\). Here it is obtained by dividing the total heat requirement for sorbent regeneration \((Q)\) by the total amount of CO\(_2\) captured \((m_{\text{CO2}})\).

\[
q_{\text{regen}} = \frac{Q}{m_{\text{CO2}}}
\]

**Enthalpy of regenerating steam \((q_{\text{steam}})\)**
The regeneration heat is provided in the form of LP steam extracted from the steam turbine (in case of coal-fired power plants and combined-cycle gas plants), through the reboiler (a heat exchanger). In case of simple cycle natural gas fired power plants, a heat recovery unit may be required. \((h_{\text{steam}})\) is the enthalpy or heat content of the steam used for solvent regeneration. Typically, the LP steam is around 300°C and 60-80 psi. From the steam-tables, the enthalpy (heat content) of such steam is found to be about 2000 kJ/ kg steam.

**Heat to electricity equivalence factor \((F_{HE})\)**
The energy penalty (extraction of LP steam) results in some loss of power generation capacity of the plant. This factor \((F_{HE})\) gives the equivalent loss of power generation capacity due to the heat requirement for solvent regeneration.

From the data obtained from the available studies (Smelster et al., 1991; Mimura et al., 1997; Bolland and Undrum, 1999; Marion et al., 2001; Hendriks, 1994), this factor has been found to lie in the range \((9, 22)\) for a new plant and \((22, 30)\) for retrofit cases. So, the nominal value (for this new plant application) has been taken as 14%.

\[
F_{HE} = 14\% \quad \text{i.e.} \quad F_{HE} = 0.14 \left(\frac{\text{kW} \cdot \text{s}}{\text{kJ}}\right) \text{Electric} \quad \frac{\text{kJ}}{\text{Heat}}
\]

So, if 10,000 kJ is the regeneration heat requirement for CO\(_2\) capture operation, then the corresponding loss in power generation capacity of the power plant is estimated as 14% of 10,000 kJ i.e. 1400 kW.s, or \((1400/3600 = )\) 0.39 kWh. It may be noted that, in case of retrofit applications, the energy penalty might be significantly higher, and \(F_{HE}\) may be around 25%.

**Blower pressure head \((\Delta P_{\text{fg}})\)**
The flue gas has to be compressed in a flue gas blower so that it can overcome the pressure drop in the absorber tower. \((\Delta P_{\text{fg}})\) is the pressure head that needs to be provided to the flue gas in the blower, and is is about 26 kPa (~3.8 psi).
Blower (fan) efficiency \( (\eta_{\text{blower}}) \)
This is the efficiency of the fan/blower to convert electrical energy input into mechanical work output. Typically, the value of blower efficiency \( (\eta_{\text{blower}}) \) is about 75%.

Solvent head \( (\Delta P_{\text{solvent}}) \)
The solvent has to flow through the absorber column (generally through packed media) countercurrent to the flue gas flowing upwards. So, some pressure loss is encountered in the absorber column and sufficient solvent head has to be provided to overcome these pressure losses. \( (\Delta P_{\text{solvent}}) \) is the pressure head to be provided to the solvent using solvent circulation pumps. A typical value is about 200 kPa (~ 30 psi).

Pump efficiency \( (\eta_{\text{pump}}) \)
This is the efficiency of the solvent circulation pumps to convert electrical energy input into mechanical energy output. Typically, the value of \( (\eta_{\text{pump}}) \) is assumed to be 75%.

CO\textsubscript{2} product pressure \( (P_{\text{CO}_2}) \)
The CO\textsubscript{2} product may have to be carried over long distances. Hence it is necessary to compress (and liquefy) it to very high pressures \( (P_{\text{CO}_2}) \), so that it maybe delivered to the required destination in liquid form and (as far as possible) without recompression facilities en route. The critical pressure for CO\textsubscript{2} is about 1070 psig. The typically reported value of final pressure to which the product CO\textsubscript{2} stream has to be pressurized using compressors, before it is transported is about 2000 psig.

Energy required for CO\textsubscript{2} compression \( (e_{\text{comp}}) \)
This is the electrical energy required (kWh per tonne CO\textsubscript{2}) to compress a unit mass of CO\textsubscript{2} product stream to the designated pressure \( (P_{\text{CO}_2}) \) expressed in psig. Compression of CO\textsubscript{2} to high pressures takes lot of energy, and is a principle contributor to the overall energy penalty of a CO\textsubscript{2} capture unit in a power plant.

CO\textsubscript{2} compression efficiency \( (\eta_{\text{comp}}) \)
This is the effective efficiency of the compressors used to compress CO\textsubscript{2} to the desirable pressure. Typically, the value of compressor efficiency \( (\eta_{\text{comp}}) \) is about 80%. It maybe noted that the energy requirement calculated from the performance equation \( (e_{\text{comp}}) \) has to be corrected by this efficiency factor in order to get the total energy required for CO\textsubscript{2} compression.

The following set of parameters are relevant only if the CO\textsubscript{2} capture system has been configured to include an auxiliary NG boiler to supply sorbent regeneration heat.

Heating value of natural gas \( (\text{NGHV}) \)
This is the high heating value (HHV, MJ/ kmole NG) of the natural gas used as fuel for the auxiliary boiler.

Density of natural gas \( (\rho_{\text{NG}}) \)
This is the density (lb/ft\textsuperscript{3}) of the natural gas used as fuel for the auxiliary boiler.

Average molecular weight of natural gas \( (\text{mw}_{\text{NG}}) \)
This is the average molecular weight (kg / kmole NG) of the natural gas used as fuel for the auxiliary boiler. This is a function of the molar composition of the natural gas.
**Flow rate of natural gas (m\(_{NG}\))**
This is the total molar flow rate (kmole NG / hr) of the natural gas used as fuel for the auxiliary boiler. It is basically a function of the total heat requirement for sorbent regeneration in the amine system.

**Auxiliary NG boiler efficiency (\(\eta_{NGB}\))**
This is the efficiency of the auxiliary boiler that uses natural gas as fuel input. It is defined as the ratio of total thermal energy (in the form of steam) delivered by the boiler divided by the total heat energy input (in the form of heating value of the natural gas input).

**Secondary steam turbine power generation efficiency (\(\eta_{ST2}\))**
This is the efficiency of the secondary steam turbine added along with the auxiliary NG boiler to generate electrical power. It may be defined as the ratio of electrical energy generated (MW\(_{ST2}\)) by the steam turbine divided by the total thermal energy (in the form of steam) input from the auxiliary NG boiler. It is assumed that the rest of the thermal energy is contained in the LP exhaust steam from the turbine, which is sent to the reboiler for sorbent regeneration.

### 3.4. Performance Equations

The performance equations define the functional relationships among various key performance parameters. They have been derived as multivariate linear regression equations from the data obtained from the process simulation model runs.

\[
(L/G) = \exp \left( -1.4352 + 0.1239y_{CO2} + 3.4863\phi_{\text{lean}} + 0.0174\eta_{CO2} - 0.0397C + 0.0027T_{fg,in} \right) \quad \text{[adj. } R^2 = 0.92]\]

\[
(Q/L) = \exp \left( 2.5919 - 0.0059y_{CO2} - 6.3536\phi_{\text{lean}} + 0.0259C - 0.0015\eta_{CO2} \right) \quad \text{[adj. } R^2 = 0.96]\]

\[
(T_{fg,out}) = 41.15 + 0.062T_{fg,in} + 1.307y_{CO2} - 18.872\phi_{\text{lean}} + 0.270C \quad \text{[adj. } R^2 = 0.92]\]

\[
(mw_{\text{lean}}) = 16.907 + 2.333\phi_{\text{lean}} + 0.204C \quad \text{[adj. } R^2 = 0.95]\]

\[
(e_{\text{comp}}) = -51.632 + 19.207\ln(P_{CO2} + 14.7) \quad \text{[adj. } R^2 > 0.99]\]

where,

- \(L\) = total sorbent flow rate (kmole/ hr)
- \(G\) = total inlet flue gas flow rate (kmole/ hr)
- \((L/G)\) = total liquid (sorbent) applied per unit flue gas flow rate in absorber (ratio of molar flow rates)
- \(Q\) = total sorbent regeneration heat requirement (GJ/ hr)
- \((Q/L)\) = total regeneration heat supplied per unit of sorbent flow (MJ/ kmole)
\[ y_{CO_2} = \text{CO}_2 \text{ concentration in the inlet flue gas (mole \%)} \]
\[ \phi_{\text{lean}} = \text{lean sorbent CO}_2 \text{ loading (mole CO}_2/ \text{mole MEA)} \]
\[ \eta_{CO_2} = \text{CO}_2 \text{ capture efficiency (\%)} \]
\[ C = \text{MEA concentration in the sorbent (wt \%)} \]
\[ T_{fg, in} = \text{Temperature of the flue gas entering the CO}_2 \text{ absorber (deg C)} \]
\[ T_{fg, out} = \text{Temperature of the flue gas leaving the CO}_2 \text{ absorber (deg C)} \]
\[ mw_{\text{lean}} = \text{Average molecular weight of the lean sorbent (kg/ kmole sorbent)} \]
\[ e_{\text{comp}} = \text{Unit energy requirement for CO}_2 \text{ compression (kWh/ tonne CO}_2 \text{)} \]
\[ P_{CO_2} = \text{Desired CO}_2 \text{ product pressure (psig)} \]

3.5. Model Outputs

The model has been built in Analytica, which specializes in propagation of uncertainties. The key outputs of the amine system performance model include:

- **MEA requirement.** This depends mainly on the mass flow rate of CO\(_2\) in the flue gas, the desired CO\(_2\) capture efficiency, MEA concentration, and CO\(_2\) loadings in the solvent. Depending on the level of impurities in the flue gas, there is some loss of solvent. If the power plant does not have emission controls for SO\(_x\) and NO\(_x\), the cost imposed due to amine loss may be significant.

- **Energy requirement.** Heat for solvent regeneration is derived from low-pressure steam available in the power plant, which decreases power generation efficiency. Additional electrical energy is required for CO\(_2\) product compression, solvent circulation, and other system requirements. The energy requirement is one of the most important results, as it dictates the net size of the power plant, and hence the net cost of power generation and CO\(_2\) avoidance.

The following material and energy flows are estimated using the above stated inputs

- Total quantity of CO\(_2\) captured,
  \[ m_{CO_2} \text{ (tonne/hr)} = \dot{n}_{CO_2} \times (\text{MolWt})_{CO_2} \]
  \[ = \eta_{CO_2} \times n_{CO_2, \text{inlet}} \times (\text{MolWt})_{CO_2} \]
  where
  \[ \dot{n}_{CO_2} = \text{Total moles of CO}_2 \text{ captured (kmole CO}_2/ \text{hr)} \]
  \[ n_{CO_2, \text{inlet}} = \text{Molar flow rate of CO}_2 \text{ in the inlet flue gas (kmole CO}_2/\text{hr)} \]
  \[ (\text{MolWt})_{CO_2} = \text{Molecular weight of CO}_2 = 0.044 \text{ tonne/ kmole CO}_2 \]

- Net loss of MEA = MEA makeup requirement = \[ m_{\text{MEA, makeup}} \]
Estimation of total sorbent circulation flow rate: From the performance equations, we find

\[ \frac{L}{G} = f(y_{CO_2}, \phi_{lean}, \eta_{CO_2}, C, T_{fg,in}) \text{, and } L = G \times (L/G) \]

Including the makeup MEA quantity gives the total sorbent flow rate (m\(^3\)/hr)

\[
L_{tot,v} = \left\{ G \times (L/G) \times m_{\text{lean}} + \dot{m}_{\text{MEA,makeup}} \times (100/C) \right\} \times \rho_{\text{sorbent}}
\]

- Waste generated from reclaimer:

\[
m_{\text{waste}} = \left( \text{MEA lost due to acid gases} + \text{Total qty of acid gases removed} \right) + \left( \text{MEA loss due to oxidation} \right)
\]

Considering \( f_{w,\text{waste}} \) as the water content (% w/w) in the waste, the actual mass flow rate of waste is obtained as:

\[
M_{\text{waste,total}} = m_{\text{waste}} / f_{w,\text{waste}} \text{ kg/hr}
\]

Typically, the reclaimer waste contains about 40% water.

- Activated carbon consumption

\[
m_{\text{act-C}} = \dot{m}_{\text{act-C}} \times \dot{m}_{\text{CO}_2} \text{ kg act-C/hr}
\]

- Caustic Consumption in Reclaimer

\[
m_{\text{Caustic}} = \dot{m}_{\text{NaOH}} \times \dot{m}_{\text{CO}_2} \text{ kg NaOH/hr}
\]

- Process Water requirement

Unit process water makeup = \( \dot{n}_{\text{pw}} \) (tonne/hr)/MW(net)

Typically, the value of \( \dot{n}_{\text{pw}} \) is about 0.114 tonne/hr per MW(net) (Smelster et al., 1991). Therefore, the process water requirement is:

\[
(M_{\text{pw}}) = \dot{n}_{\text{pw}} \times MW_{\text{net}} \text{ tonne/hr}
\]

- Cooling water requirement
If there is a direct contact cooler installed, the required flow rate of cooling water is estimated based on the following assumptions:

Specific heat of water, \( S_{Hw} = 4.2 \text{ kJ/kg } ^\circ C \)
Specific heat of flue gas = \( S_{Hfg} \) (Generally, this is around 1.2 kJ/kg \( ^\circ C \))
Temperature rise in the cooling water (once through system) = \( \Delta T_w \)
Drop in flue gas temperature = \( \Delta T_{fg} = (T_{fg,i} - T_{fg}) ^\circ F \)

where,

\( T_{fg,i} = \) Temperature of flue gas entering the direct contact cooler
\( T_{fg} = \) Temperature of flue gas exiting the direct contact cooler

Mass flow rate of flue gas = \( m_{fg} \text{ tonne/hr} \)

So, the required cooling water flow rate,

\[ M_{cw} = m_{fg} * (\frac{\Delta T_{fg}}{\Delta T_w}) * \left( \frac{S_{Hfg}}{S_{Hw}} \right) \text{ tonne/hr} \]

Therefore, the total water requirement is:

\[ (M_w) = \text{Process water (} M_{pw} \text{) + Cooling water (} M_{cw} \text{)} \]

**Steam requirement**

LP steam is extracted from the power plant steam turbine (or secondary steam turbine) in order to provide the sorbent regeneration heat in the reboiler. Based on the regeneration heat requirement and enthalpy of regeneration steam, the flow rate of steam may be estimated as follows.

From the performance equations,

\[ (Q/L) = f(y_{CO2}, \phi_{lean}, C) \]

Total regeneration heat requirement,

\[ Q \text{ (MJ/hr)} = (Q/L)*L \]

Mass flow rate of steam,

\[ m_{steam} \text{ (tonne/hr)} = \frac{Q}{\phi_{steam}} \]

The equivalent energy penalty due to regeneration steam requirement is \( E_{regen} \). Depending upon the \( CO_2 \) capture system configuration (source of regeneration steam supply), \( E_{regen} \) has to be estimated in two different ways.

**August 2011 Update:** In the most recent version of the IECM, the steam use at the reboiler is calculated using a new set of formulas. Please see "IECM Technical Documentation Updates, Vol. I to V (Final Report to USDOE/NETL from Carnegie Mellon University)", for details.

1. In case of steam extraction from the base plant steam cycle (derating)

\[ E_{regen} = Q*F_{HE} \]

2. In case of steam supplied from an auxiliary NG boiler,

\[ E_{regen} = - E_{ST2} = - (m_{NG}*NG*HV*\eta_{NGB}*\eta_{ST}) \]
It maybe noted that in the case of auxiliary NG boiler, the energy penalty term is negative, implying that there is an increase in the net power generation of the plant.

Total energy penalty of CO\textsubscript{2} capture system is:

$$E_{\text{CO}_2}^{\text{tot}} = E_{\text{regen}} + E_{\text{pumping}} + E_{\text{compr}}$$

where,

- $E_{\text{regen}} = $ as explained in (9)
- $E_{\text{pumping}} = E_{\text{blower}} + E_{\text{pump}}$

$$E_{\text{blower}} \text{ (hp)} = \frac{144Q_{fg} \Delta P_{fg}}{33000 \cdot \eta_{\text{blower}}}$$

where $Q_{fg}$ and $\Delta P_{fg}$ are expressed in ft\textsuperscript{3}/min and psi, respectively,

$$E_{\text{pump}} \text{ (hp)} = \frac{2Q_{\text{solvent}} \Delta P_{\text{solvent}}}{1714 \cdot \eta_{\text{pump}}}$$

where $Q_{\text{solvent}}$ and $\Delta P_{\text{solvent}}$ are expressed in gal/min and psi, respectively and,

$$E_{\text{compr}} = e_{\text{comp}} \frac{m_{\text{CO}_2}}{\eta_{\text{comp}}}$$

### 3.6. Characterization of Uncertainties

One of the distinguishing features of this modeling effort is a probabilistic capability that allows model inputs to be represented by probability distributions rather than single deterministic values. Uncertainties in these parameters reflect the ranges of values reported in the literature, the evolving nature of the technology, and practical considerations in running such plants. Table 4 lists the uncertainty distributions developed for performance model parameters based on the current literature on amine-based (MEA) systems. These distributions reflect both uncertainty and variability in system designs.

**Table 4. Amine System Performance Model Parameters and Uncertainties**

<table>
<thead>
<tr>
<th>Performance Parameter</th>
<th>Units</th>
<th>Data (Range)</th>
<th>Nominal Value</th>
<th>Unc. Representation (Distribution Function)</th>
</tr>
</thead>
<tbody>
<tr>
<td>CO\textsubscript{2} removal efficiency</td>
<td>%</td>
<td>Mostly 90</td>
<td>90</td>
<td>-</td>
</tr>
<tr>
<td>SO\textsubscript{2} removal efficiency</td>
<td>%</td>
<td>Almost 100</td>
<td>99.5</td>
<td>Uniform(99,100)</td>
</tr>
<tr>
<td>NO\textsubscript{2} removal efficiency</td>
<td>%</td>
<td>20-30</td>
<td>25</td>
<td>Uniform(20,30)</td>
</tr>
<tr>
<td>HCl removal efficiency</td>
<td>%</td>
<td>90-95</td>
<td>95</td>
<td>Uniform(90,95)</td>
</tr>
<tr>
<td>Particulate removal eff.</td>
<td>%</td>
<td>50</td>
<td>50</td>
<td>Uniform(40,60)</td>
</tr>
<tr>
<td>MEA concentration</td>
<td>wt%</td>
<td>15-50</td>
<td>30</td>
<td>-</td>
</tr>
<tr>
<td>Lean solvent CO\textsubscript{2} loading</td>
<td>mol CO\textsubscript{2}/mol MEA</td>
<td>0.15-0.30</td>
<td>0.22</td>
<td>Triangular(0.17,0.22,0.25)</td>
</tr>
<tr>
<td>Nominal MEA make-up</td>
<td>kg MEA/tonne CO\textsubscript{2}</td>
<td>0.5-3.1</td>
<td>1.5</td>
<td>Triangular(0.5,1.5,3.1)</td>
</tr>
<tr>
<td>MEA loss (SO\textsubscript{2})</td>
<td>mol MEA/mol SO\textsubscript{2}</td>
<td>2</td>
<td>2</td>
<td>-</td>
</tr>
<tr>
<td>MEA loss (NO\textsubscript{2})</td>
<td>mol MEA/mol NO\textsubscript{2}</td>
<td>2</td>
<td>2</td>
<td>-</td>
</tr>
<tr>
<td>MEA loss (HCl)</td>
<td>mol MEA/mol HCl</td>
<td>1</td>
<td>1</td>
<td>-</td>
</tr>
</tbody>
</table>
### Table

<table>
<thead>
<tr>
<th>Parameter</th>
<th>Unit</th>
<th>Value</th>
<th>Value</th>
<th>Distribution</th>
</tr>
</thead>
<tbody>
<tr>
<td>MEA loss (exhaust gas)</td>
<td>ppm</td>
<td>1-4</td>
<td>2</td>
<td>Uniform (1,4)</td>
</tr>
<tr>
<td>NH₃ generation</td>
<td>mol NH₃/mol MEA oxidized</td>
<td>1</td>
<td>1</td>
<td>-</td>
</tr>
<tr>
<td>Caustic consumption in MEA reclaimer</td>
<td>kg NaOH/tonne CO₂</td>
<td>0.13</td>
<td>0.13</td>
<td>-</td>
</tr>
<tr>
<td>Activated carbon use</td>
<td>kg C/tonne CO₂</td>
<td>0.075</td>
<td>0.075</td>
<td>-</td>
</tr>
<tr>
<td>Cooling water makeup</td>
<td>m³/tonne CO₂</td>
<td>0.5-1.8</td>
<td>0.8</td>
<td>Triangular (0.5, 0.8, 1.8)</td>
</tr>
<tr>
<td>Solvent pumping head</td>
<td>kPa</td>
<td>35-250</td>
<td>207</td>
<td>Triangular (150, 207, 250)</td>
</tr>
<tr>
<td>Pump efficiency</td>
<td>%</td>
<td>70-75</td>
<td>75</td>
<td>Uniform (70,75)</td>
</tr>
<tr>
<td>Gas-phase pressure drop</td>
<td>kPa</td>
<td>14-30</td>
<td>26</td>
<td>Triangular (14,26,30)</td>
</tr>
<tr>
<td>Fan efficiency</td>
<td>%</td>
<td>70-75</td>
<td>75</td>
<td>Uniform (70,75)</td>
</tr>
<tr>
<td>Equiv. elec. requirement</td>
<td>% regeneration heat</td>
<td>9-19</td>
<td>14⁺</td>
<td>Uniform (9,19)</td>
</tr>
<tr>
<td>CO₂ product purity</td>
<td>wt%</td>
<td>99-99.8</td>
<td>99.5</td>
<td>Uniform (99,99.8)</td>
</tr>
<tr>
<td>CO₂ product pressure</td>
<td>MPa</td>
<td>5.86-15.16</td>
<td>13.79</td>
<td>Triangular (5.86,13.79,15.16)</td>
</tr>
<tr>
<td>Compressor efficiency</td>
<td>%</td>
<td>75-85</td>
<td>80</td>
<td>Uniform (75,85)</td>
</tr>
</tbody>
</table>

⁺ For retrofit applications, nominal value is 25.

### 4. COST MODEL DEVELOPMENT

The CO₂ capture and sequestration system cost model is directly linked to the performance model. The cost model follows the framework used in the IECM to ensure consistency in economic calculations. There are four types of cost calculated by this model based on the available data (Smelster et al., 1991; Hendriks, 1994; Leci, 1996; Chapel et al., 1999; Simbeck, 1999; Desideri and Paolucci, 1999; Jeremy and Herzog, 2000).

#### 4.1. Capital Cost

The total capital requirement (TCR) of a system is calculated as the sum of direct equipment costs (which depend on one or more performance variables that determine the size or capacity of the component), plus various indirect costs that are estimated as fractions of the total direct cost following the EPRI cost estimating guidelines (TAG, 1993).

The capital cost model is based on the cost and flow rate information obtained from Fluor Daniel Inc (Fluor Daniel, 1998). It is assumed that there are multiple trains installed to perform the CO₂ capture operation. Based on the same source, the maximum train size has been assumed to be 5000 tonnes per day of CO₂. Based on the actual CO₂ capture rate (nCO₂) the minimum number of trains required to be installed (Nₘᵢₙ) is determined. Different equipments have different maximum capacity limits. So, (Eₙ,i) defines the number of equipments required per train.

Eₙ,i: Each train consists of the following pieces of equipment:
- Direct contact cooler (DCC), flue gas blower, absorber, heat exchanger, regenerator, steam extractor, MEA reclaimer - 1 per each train
- Pumps - 2 per each train
- Reboilers - 4 per each train

Special cases:
1. Each train need not have a separate installation of the solvent processing area, CO₂ transport facility and CO₂ disposal facility, and they will be installed for the whole CO₂ capture unit. Hence “Eₙ” in this case, may be considered as (1/Nᵢ) per train, where Nᵢ is the total number of trains installed.

2. In case of CO₂ compressors, which have higher capacity (~ 7200 tpd CO₂), the number of compressors required is calculated accordingly. If Nᵢ is the total number of CO₂ compressors installed, then the number of compressors installed per train may be stated as (Nᵢ/Nᵢ).

Different components of this system (Absorber, Regenerator, Flue gas blower etc.) are scaled, based on the flow rate of the material being handled by that particular device, using 0.6 power law e.g., the cost of absorber and flue gas blower is scaled on the basis of flue gas flow rate entering the CO₂ system. The data obtained from the Fluor Daniel report serve as reference numbers for this scaling exercise.

Actual value of scaling parameter per train (X) is calculated by dividing the magnitude of the scaling parameter (obtained from the performance model) by the minimum number of equipments required (i.e. product of minimum number of trains required and minimum number of equipments per train). e.g. if V is the value of a parameter, then X is given as

\[ X = \frac{V}{E_{n,ibt} \cdot N_{min}} \]

So, different process areas using the same scaling parameter may have different value of X, depending upon the value of Eₙ.

Each process area has a reference cost (Cᵣₑᶠ) based on the source sited before, and the corresponding value of the scaling parameter (Xᵣₑᶠ). The cost of the equipment is calculated using the reference values and the actual value of scaling parameter (X), based on the 6/10th rule which is commonly used in chemical engineering costing. For example, in case of a particular process area (say, area 10), we have the following cost:

\[ C_{10,ref} = \text{Cost of equipment (area 10)} \]

\[ \text{Scaling parameter} = X_{10,ref} \]

From the performance model, we have: the total quantity of the scaling parameter, Y. Now, as discussed above,

\[ N_{min} = \text{Minimum number of trains} \]
\[ E_{n,10} = \text{Number of equipment (10) per train} \]
\[ Z_{10, min} = N_{min} \times E_{n,10} \]
\[ Z_{10} = N \times E_{n,10} \]

where,

Nᵢ is the actual number of trains installed (including spares)

So, the actual flue gas flow rate per train,
Therefore, the actual capital cost of absorber in this case may be estimated as

\[ C_{10} = C_{10, \text{ref}} \cdot \left( \frac{X_{10}}{X_{10, \text{ref}}} \right)^{0.6} \]

Once the cost of a particular equipment is calculated \((C_{10})\), it needs to be multiplied by the total number of equipments installed \((Z_{10})\) in order to get the total cost of installation for that process area \((10)\).

Similarly, in case of other process areas some physical quantity can be identified (e.g., flue gas flow rate, solvent flow rate, CO\(_2\) product flow rate, CO\(_2\) compression energy requirement, steam flow rate, makeup MEA flow rate etc.) that may be used for scaling of the capital cost.

The direct capital cost (process facilities) of CO\(_2\) capture and separation system consists of the following cost items

**Direct contact cooler**: In case of coal-fired power plant applications that have a wet FGD (flue gas desulfurization) unit upstream of the amine system, the wet scrubber helps in substantial cooling of the flue gases, and additional cooler may not be required. In case of gas-fired power plants or majority of coal-fired power plants that do not have wet scrubbers for SO\(_2\) removal, a direct contact cooler has to be installed to bring down the temperature of the flue gas stream to acceptable levels. A direct contact cooler is essentially a large vessel where the incoming hot flue gas is made to contact with the cooling water. The size of this unit is mainly a function of the volumetric flow rate of the flue gas, which in turn depends upon the temperature and pressure conditions of the flue gas stream. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.

\[ C_{\text{dec}} = C_{\text{dec, ref}} \cdot \left( \frac{V_{fg}}{V_{fg, \text{ref}}} \cdot \frac{T_{fg}}{T_{fg, \text{ref}}} \right)^{0.6} \]

**Flue gas blower**: The cooled flue gas is pressurized using a blower before it enters the absorber. The size (and the cost) of the blower is again a function of the volumetric flow rate of the flue gas as it enters the blower. So, the cost maybe estimated using as above

\[ C_{\text{blower}} = C_{\text{blower, ref}} \cdot \left( \frac{V_{fg,1}}{V_{fg,1, \text{ref}}} \cdot \frac{T_{fg,1}}{T_{fg,1, \text{ref}}} \right)^{0.6} \]

**Absorber**: This is the vessel where the flue gas is made to contact with the MEA-based solvent, and some of the CO\(_2\) from the flue gas gets dissolved in the solvent. Again, the size of this unit is mainly a function of the volumetric flow rate of the flue gas, which in turn depends upon the temperature and pressure conditions of the flue gas stream, as it enters this vessel. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.
\[ C_{\text{absorber}} = C_{\text{absorber, ref}} \cdot \left( \frac{V_{\text{fg, in}}}{V_{\text{fg, in, ref}}} \cdot \frac{T_{\text{fg, in}}}{T_{\text{fg, in, ref}}} \right)^{0.6} \]

**Rich/lean cross heat exchanger**: The rich (CO\(_2\)-loaded) and lean (regenerated) solvent streams are passed through this cross heat exchanger, where the rich solvent gets heated and the lean solvent gets cooled. So, the size (and cost) of this unit is mainly a function of the volumetric solvent flow rate in the absorber. It is assumed that this volumetric flow rate is fairly constant in the range of temperature and pressure conditions found in this system. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.

\[ C_{\text{crossHEX}} = C_{\text{crossHEX, ref}} \cdot \left( \frac{V_{\text{solvent}}}{V_{\text{solvent, ref}}} \right)^{0.6} \]

**Regenerator**: This is the column where the CO\(_2\)-loaded solvent is regenerated with the application of heat. Solvent flow rate is the main physical quantity that decides the size (and cost) of this unit, for a given residence time (which is a function of many parameters including the solvent concentration, desired CO\(_2\) capture efficiency, etc.). So, the cost maybe estimated using as above

\[ C_{\text{regenerator}} = C_{\text{regenerator, ref}} \cdot \left( \frac{V_{\text{solvent}}}{V_{\text{solvent, ref}}} \right)^{0.6} \]

**Reboiler**: The regenerator is connected with a reboiler, which is basically a heat exchanger where low-pressure steam extracted from the power plant is used to heat the loaded solvent. So, the size (and cost) of this unit is a function of mainly the flow rate of the solvent as well as the flow rate of steam. The actual cost of the unit is estimated on the basis of the cost information available for a particular reference case study using 0.6 power law for scaling purposes.

\[ C_{\text{reboiler}} = C_{\text{reboiler, ref}} \cdot \left( \frac{V_{\text{solvent}}}{V_{\text{solvent, ref}}} \cdot \frac{M_{\text{steam}}}{M_{\text{steam, ref}}} \right)^{0.6} \]

It maybe noted that the ratio of mass flow rates of LP steam (\(M_{\text{steam}}/ M_{\text{steam, ref}}\)) has been used in place of the ratio of volumetric flow rates of LP steam, assuming that the temperature and pressure conditions of the LP steam in both cases (actual and reference) are almost identical.

**Steam extractor**: Steam extractors are installed to take out LP/IPsteam from the steam turbines in the power plant. The size (and the cost) of the steam extractor is assumed to be a function of the steam flow rate.

\[ C_{\text{steam, extractor}} = C_{\text{steam, extractor, ref}} \cdot \left( \frac{M_{\text{steam}}}{M_{\text{steam, ref}}} \right)^{0.6} \]

This cost item is included if the CO\(_2\) capture system is configured to make use of steam extracted from the steam cycle of the base plant. Alternatively, an auxiliary NG boiler
and a secondary steam turbine maybe used, and the next two cost items (8 and 9) are included in its place.

**Auxiliary boiler:** The cost of the NG boiler is estimated on the basis of the (no reheat) steam flow rate generated from the boiler. The following cost estimation formula was reported by Dale Simbeck

\[ C_{\text{NG, boiler}} = \$15 \times \text{(steam flow rate expressed in lb/hr)} \]

Since the steam flow rate \( m_{\text{steam}} \) was estimated as tonnes/hr, the following expression maybe obtained after accounting for the unit conversions

\[ C_{\text{NG, boiler}} = \$33000 \times (m_{\text{steam}}) \]

**Secondary steam turbine:** The cost of the secondary steam turbine is estimated on the basis of the electrical power generated from this new turbine. The following cost estimation formula was reported by Dale Simbeck

\[ C_{\text{ST2}} = \$300 \times \text{(power generation expressed in kWe)} \]

Since the power generation \( E_{\text{ST2}} \) was estimated as MWe, the following expression maybe obtained after accounting for the unit conversions

\[ C_{\text{ST2}} = \$300000 \times (E_{\text{ST2}}) \]

**MEA reclaimer:** In order to avoid accumulation of the heat stable salts in the solvent stream and to recover some of the lost MEA solvent, a part of the solvent stream is periodically distilled in this vessel. Addition of caustic helps in freeing of some of the MEA. The amount of MEA makeup required, maybe taken as an indicative of the amount of heat stable salts formed and the quantity of solvent to be distilled in the reclaimer. So, the mass flow rate of makeup MEA requirement is used as a scaling parameter to estimate the cost of this unit, based on a reference study.

\[ C_{\text{MEA, reclaimer}} = C_{\text{MEA, reclaimer, ref}} \cdot \left( \frac{M_{\text{MEA, makeup}}}{M_{\text{MEA, makeup, ref}}} \right)^{0.6} \]

**Solvent processing area:** The solvent processing area primarily consists of solvent cooler, MEA storage tank, and a mixer. It also consists of an activated carbon bed filter that adsorbs impurities (degradation products of MEA) from the solvent stream. So, the size (and cost) of this unit (together) will be a function of the total solvent flow rate, and maybe estimated as follows

\[ C_{\text{solvent, proc}} = C_{\text{solvent, proc, ref}} \cdot \left( \frac{V_{\text{solvent}}}{V_{\text{solvent, ref}}} \right)^{0.6} \]

**CO₂ drying and compression unit:** The multi-stage compression unit with inter-stage cooling and drying yields a final CO₂ product at the specified pressure (about 2000 psig) that contains moisture and other impurities (e.g. N₂) at acceptable levels. Obviously, the size (and cost) of this unit will be a function of the CO₂ product flow rate, and maybe estimated as follows
\[
C_{CO_2\_compr} = C_{CO_2\_compr\_ref} \cdot \left( \frac{M_{CO_2}}{M_{CO_2\_ref}} \right)^{0.6}
\]

The sum of all these individual process area equipment costs is termed as process facilities cost (PFC). The various indirect costs are then estimated as fractions of the total direct cost (PFC) following the EPRI cost estimating guidelines (TAG, 1993).

Table 5 lists the elements of total capital cost. Because of data limitations some of the indirect cost factors are estimated based on other technologies.

<table>
<thead>
<tr>
<th>Capital Cost Elements</th>
<th>Value</th>
</tr>
</thead>
<tbody>
<tr>
<td>A</td>
<td>Process Area Equipment Costs</td>
</tr>
<tr>
<td>B</td>
<td>Total Process Facilities Cost (PFC)</td>
</tr>
<tr>
<td>C</td>
<td>Engineering and Home Office</td>
</tr>
<tr>
<td>D</td>
<td>General Facilities</td>
</tr>
<tr>
<td>E</td>
<td>Project Contingency</td>
</tr>
<tr>
<td>F</td>
<td>Process Contingency</td>
</tr>
<tr>
<td>G</td>
<td>Total Plant Cost (TPC) = sum of above</td>
</tr>
<tr>
<td>H</td>
<td>Interest Costs During Constr.</td>
</tr>
<tr>
<td>I</td>
<td>Royalty Fees</td>
</tr>
<tr>
<td>J</td>
<td>Pre-production (Fixed O&amp;M)</td>
</tr>
<tr>
<td>K</td>
<td>Pre-production (Variable O&amp;M Cost)</td>
</tr>
<tr>
<td>L</td>
<td>Inventory (startup) Cost</td>
</tr>
<tr>
<td>M</td>
<td>Total Capital Requirement (TCR)°</td>
</tr>
</tbody>
</table>

4.2 O&M Cost

The major operating and maintenance (O&M) cost consists of some fixed costs and some variable cost elements as listed in Table 6.

<table>
<thead>
<tr>
<th>O&amp;M Cost Elements</th>
<th>Typical Value</th>
</tr>
</thead>
<tbody>
<tr>
<td><strong>Fixed O&amp;M Costs</strong></td>
<td></td>
</tr>
<tr>
<td>Total Maintenance Cost</td>
<td>2.5% TPC</td>
</tr>
<tr>
<td>Maintenance Cost Allocated to Labor (f_maintlab)</td>
<td>40% of total maint. cost</td>
</tr>
<tr>
<td>Admin. &amp; Support Labor Cost (f_admin)</td>
<td>30% of total labor cost</td>
</tr>
<tr>
<td>Operating Labor (N_labor)</td>
<td>2 jobs/shift</td>
</tr>
<tr>
<td><strong>Variable O&amp;M Costs</strong></td>
<td></td>
</tr>
<tr>
<td>Reagent (MEA) Cost</td>
<td>$1250/ mton</td>
</tr>
<tr>
<td>Water Cost</td>
<td>$0.8/ 1000 gallon</td>
</tr>
<tr>
<td>CO₂ Transport Cost</td>
<td>$0.04/ mton CO₂ km</td>
</tr>
<tr>
<td>CO₂ Storage/Disposal Cost</td>
<td>$5/ mton CO₂</td>
</tr>
<tr>
<td>---------------------------------</td>
<td>-------------</td>
</tr>
<tr>
<td>Solid Waste Disposal Cost</td>
<td>$175/ mton waste</td>
</tr>
</tbody>
</table>

The *fixed O&M* (FOM) costs include the costs of maintenance (materials and labor) and labor (operating labor, administrative and support labor). These are estimated on the annual basis ($M/yr). The mathematical model for the fixed cost is as follows

\[
FOM = FOM_{labor} + FOM_{maint} + FOM_{admin}
\]

\[
FOM_{labor} = \text{labor} \times N_{labor} \times 40(\text{hrs/week}) \times 52(\text{weeks/yr})
\]

\[
FOM_{maint} = \sum_i (f_{\text{main}}) \times TPC_i \quad \text{where } i = \text{process area}
\]

\[
FOM_{admin} = f_{\text{admin}} \times (FOM_{labor} + f_{\text{maint}} \times FOM_{main})
\]

The *variable O&M* (VOM) costs include:

**Cost of MEA reagent** (VOM\textsubscript{MEA}): The makeup MEA requirement estimated in the performance model is transformed into dollar amount by using the unit cost of MEA, which is user controlled cost input variable.

\[
VOM_{\text{MEA}} = M_{\text{MEA, makeup}} \times UC_{\text{MEA}} \times \text{HPY}
\]

where, \(UC_{\text{MEA}}\) is the unit cost of MEA.

**Cost of inhibitor** (VOM\textsubscript{inhibitor}): Addition of inhibitor makes it possible to use higher concentrations of MEA solvent in the system with minimal corrosion problems. Inhibitors are special compounds that come at a cost premium. The cost of inhibitor is estimated as 20% of the cost of MEA.

\[
VOM_{\text{inhibitor}} = 0.2 \times VOM_{\text{MEA}}
\]

**Cost of other reagents** (VOM\textsubscript{reagents}): The cost of other reagents, such as, caustic and activated carbon are also calculated from their physical quantities estimated in the performance model and the unit costs of these reagents.

\[
VOM_{\text{reagents}} = VOM_{\text{Caustic}} + VOM_{\text{act-C}}
\]

\[
= \{(m_{\text{Caustic}} \times UC_{\text{Caustic}}) + (m_{\text{act-C}} \times UC_{\text{act-C}})\} \times \text{HPY}
\]

where, \(UC_{\text{Caustic}}\) and \(UC_{\text{act-C}}\) are the unit costs of the reagents caustic and activated carbon, respectively.

**Cost of waste disposal** (VOM\textsubscript{waste}): Another important variable operating cost item is the cost incurred in proper disposal of the spent sorbent i.e. the reclamer waste, again the quantity estimated in the performance model.

\[
VOM_{\text{waste}} = M_{\text{waste,total}} \times UC_{\text{waste}} \times \text{HPY}
\]

where, \(UC_{\text{waste}}\) is the unit cost of waste disposal for the reclamer waste.

**Cost of CO₂ transport** (VOM\textsubscript{transport}): Transportation of CO₂ product is assumed to take place via pipelines. The cost of CO₂ transport is estimated on the basis of two user specified...
parameters, viz., transportation distance (TD, in km) and unit cost of transport ($/km mton CO\(_2\)), and CO\(_2\) product flow rate (calculated result from performance model).

\[ VOM_{\text{transport}} = M_{\text{CO}_2} \times UC_{\text{transport}} \times TD \times HPY \]

**Cost of CO\(_2\) disposal** (VOM\(_{\text{disposal}}\)): Depending upon the method of CO\(_2\) disposal or storage, either there may be some revenue generated (Enhanced Oil Recovery, Coal Bed Methane) which may be treated as a “negative cost”, or additional cost (all other disposal methods). The total cost or revenue of CO\(_2\) disposal/ storage is estimated from the unit cost and CO\(_2\) product flow rate (UC\(_{\text{disp}}\)).

\[ VOM_{\text{disposal}} = M_{\text{CO}_2} \times UC_{\text{disp}} \times HPY \]

**Cost of energy** (VOM\(_{\text{energy}}\)): By default, the energy costs are handled internally in the model by de-rating the overall power plant based on the calculated power requirement. This increases the cost per net kilowatt-hour delivered by the plant. The CO\(_2\) capture unit is charged for the total electricity production foregone (energy penalty) because of capture and compression of CO\(_2\) from the flue gas, and the base plant is credited for the same. The unit cost of electricity (COE\(_{\text{noccl}}\)) is estimated by the base plant module, or maybe overridden by a user-specified value when this energy is supplied from an external source (in that case, no credit given to the base plant). Since energy cost is one of the biggest O&M cost items for CO\(_2\) unit, the way in which it is accounted for (internal de-rating or external provision) becomes very crucial while calculating the mitigation cost.

\[ VOM_{\text{energy}} = E_{\text{CO}_2,\text{tot}} \times HPY \times COE_{\text{noccl}} \]

Alternatively, when regeneration steam is provided by an auxiliary NG boiler, the cost of energy maybe estimated from the total annualized cost of the new boiler and secondary steam turbine, which takes into account their capital cost requirement and cost of natural gas fuel.

**Cost of water** (VOM\(_{\text{water}}\)): Water is mainly required for cooling and also as process makeup. Generally this is a minor cost item in the overall plant operation, but it is included over here for the sake of completeness. Also, it maybe noted that the unit cost of water (UC\(_{\text{water}}\)) may vary depending upon the location of the power plant.

\[ VOM_{\text{water}} = M_w \times UC_{\text{water}} \times HPY \]

So, the total variable O&M (VOM, $/yr) cost is obtained by adding all these costs

\[ VOM = VOM_{\text{MEA}} + VOM_{\text{reagents}} + VOM_{\text{waste}} + VOM_{\text{transport}} + VOM_{\text{disposal}} + VOM_{\text{energy}} + VOM_{\text{water}} \]

Finally, the total annual O&M cost (TOM, $/yr) maybe obtained as

\[ TOM = FOM + VOM \]

### 4.3 Incremental Cost of Electricity

Once the total capital cost requirement and the total O&M costs are known, the total annualized cost of the power plant may be estimated as follows:

\[ \text{Total annualized cost, TRR ($/yr)} = TCR \times CRF + TOM \]

Where, TCR = Total capital requirement of the power plant ($), and
CRF = Capital recovery factor (%)
The IECM framework calculates the cost of electricity (COE) for the overall power plant by dividing the total annualized plant cost ($/yr) by the net electricity generated (kWh/yr). Results are expressed in units of $/MWh (equivalent to mills/kWh). Two key parameters are the capital recovery factor (to amortize capital expenses), and the plant capacity factor (which determines the effective annual hours of operation at full load).

\[
\text{Cost of electricity, COE ($/MWh)} = \frac{\text{TRR}}{\text{MW}_{\text{net}} \times \text{HPY}}
\]

Where,

\[
\text{TRR} = \text{Total annualized cost ($/yr)}
\]

\[
\text{MW}_{\text{net}} = \text{Net power generation capacity (MW)}
\]

\[
\text{HPY} = \text{Annual hours of operation (hrs/yr)}
\]

So, by running two scenarios of the power plant model, one without CO\(_2\) capture unit (reference plant) and one with CO\(_2\) capture unit (CO\(_2\) capture plant), we obtain the respective capital costs, O&M costs to give the annualized costs (TRR) and finally the cost of electricity (COE) with and without CO\(_2\) capture. The addition of a CO\(_2\) capture and sequestration system increases the COE for the plant; this incremental cost of electricity is attributed to CO\(_2\) control.

### 4.4 Cost of CO\(_2\) Avoided

Many analysts like to express the cost of an environmental control system in terms of the cost per ton of pollutant removed or avoided. For energy-intensive CO\(_2\) controls there is a big difference between the cost per ton CO\(_2\) removed and the cost per ton “avoided” based on net plant capacity. Since the purpose of adding a CO\(_2\) unit is to reduce the CO\(_2\) emissions per net kWh delivered, the cost of CO\(_2\) avoidance is the economic indicator that is widely used in this field. It can be calculated as:

\[
\text{Cost of CO}_2 \text{ Avoided ($/mton)} = \frac{(S/kWh)_{\text{after}} - (S/kWh)_{\text{before}}}{(t \ CO_2 / kWh)_{\text{before}} - (t \ CO_2 / kWh)_{\text{after}}}
\]

For power plants with multi-pollutant controls the desire to quantify costs for a single pollutant sometime requires an arbitrary choice of how to charge or allocate certain costs. This is especially relevant for energy-intensive processes like CO\(_2\) capture systems.

The cost of CO\(_2\) avoidance has another interpretation in terms of the carbon-tax scenarios. Consider a scenario where every power plant is made to pay a fixed amount of tax (C-tax) that is proportional to their CO\(_2\) emissions. Now let’s have a reference plant (one that does not control its CO\(_2\) emissions) and a CO\(_2\) capture plant (one that captures, say 90% of its CO\(_2\) emissions). The reference plant will pay a much higher C-tax (almost 10 times that paid by the capture plant). So, the COE for the reference plant increases much faster as compared to the COE for the capture plant, in response to increasing levels of the C-tax. Eventually, a C-tax level maybe reached where COE for both the plants are same (see Figure 6).

It means that at this C-tax level, the power plant might be indifferent between paying C-tax for its entire CO\(_2\) emissions or incurring the cost of the CO\(_2\) capture unit. Above this particular C-tax level, the COE for the reference plant will be higher than that for the capture plant, as it is evident
from the figure. So, cost of CO$_2$ avoidance is this C-tax level, where the COE for the reference plant and capture plant become equal.

![Figure 6. Cost of electricity (COE) as a function of carbon-tax](image)

5. **UNCERTAINTY DISTRIBUTION BASED ON DATA FOR COMMERCIAL SYSTEMS (WORK IN PROGRESS)**

6. **REFERENCES**


