IECM Technical Documentation:

Amine-based Post-Combustion CO_2 Capture

Compiled in January 2019

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Amine-based Post-Combustion CO_2 Capture

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Acknowledgements

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- Anand B. Rao and Edward S. Rubin. A Technical, Economic and Environmental Assessment of Amine-based CO_2 Capture Technology for Power Plant Greenhouse Gas Control. Prepared by Carnegie Mellon University for the National Energy Technology Laboratory. Pittsburgh, PA 15213, October 2002.
- Anand B. Rao. Details of a Technical, Economic and Environmental Assessment of Amine-based CO_2 Capture Technology for Power Plant Greenhouse Gas Control. Prepared by Carnegie Mellon University for the National Energy Technology Laboratory. Pittsburgh, PA 15213, October 2002.
- Michael B. Berkenpas, Karen Kietzke, Hari Mantripragada, Sean McCoy, Edward S. Rubin, Peter L. Versteeg, and Haibo Zhai. IECM Technical Documentation Updates, Vol. IV. Prepared by Carnegie Mellon University for the National Energy Technology Laboratory. Pittsburgh, PA 15213, November 2009.

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1. Introduction

1.1. Technology Options for CO_2 Capture

A wide range of technologies currently exist for separation and capture of CO_2 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 CO_2 needs to be separated. Future power plants may be designed so as to separate out CO_2 from coal before combustion (using coal-gasification systems), or they may employ pure oxygen combustion instead of air so as to obtain a concentrated CO_2 stream for treatment. Figure 2 shows the variety of power plant fuels and technologies that affect the choice of CO_2 capture systems. In this report, post-combustion capture of CO_2 from flue gas streams of conventional power plant using amine-based absorption systems has been considered.

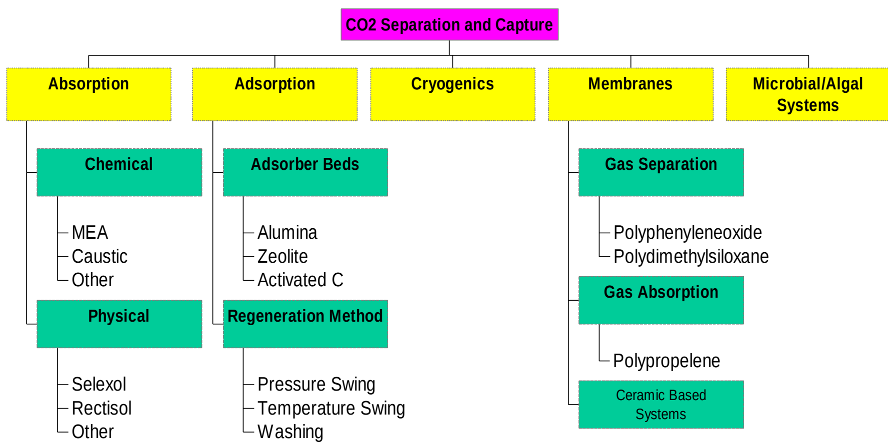


Figure 1. Technology Options for CO_2 Separation and Capture

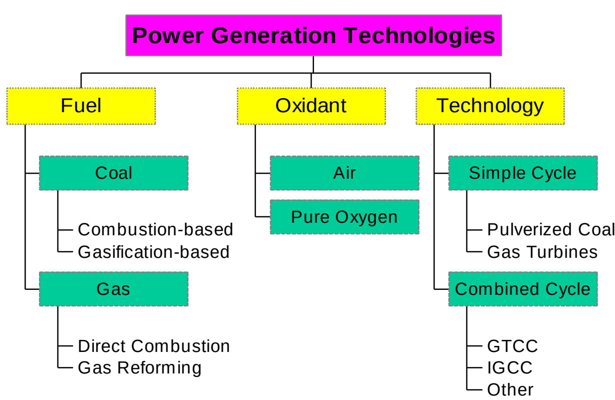


Figure 2. Technology Options for Fossil-Fuel based Power Generation

Table 1. Comparison of Technology Options for CO_2 Separation and Capture

Technology Option	**System Requirements**	**Advantages**	**Problems/ Drawbacks**
Absorption (Chemical)	Absorber and stripper sections Chemical solvent (*e.g.* MEA, HPC)	Suitable for dilute CO_2 streams (typical flue gas from power plants) Operates at ordinary T & P Commercially available, proven technology	The heat of solvent regeneration is very high Significant solvent losses due to acidic impurities in the gas stream
Absorption (Physical)	Absorber and stripper sections Physical solvent (*e.g.* Selexol)	Less energy required Solvents are less susceptible to the impurities in the gas stream	Requires high operating pressure Works better with gas streams having high CO_2 content
Adsorption	Adsorber bed(s)	Very high CO_2 removal is possible	Requires very high operating pressures Costly
Membranes	Membrane filter(s)	Upcoming, promising technology Space efficient	Requires very high operating pressures May require multiple units and recycling due to lower product purity Very costly

1.2. Post-combustion amine-based absorption of CO_2 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 CO_2 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 CO_2-emissions has received very little attention to date.

In current systems which use air for combustion, post-combustion CO_2 separation from the flue gas stream has to be carried out. Past studies have shown that amine-based CO_2 absorption systems are the most suitable for combustion-based power plants for the following reasons

- These systems are effective for dilute CO_2 streams, such as coal combustion flue gas which typically contains about 10%-12 % CO_2 by volume.

- Amine-based CO_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_2 abatement. Thus, one can anticipate future benefits from technology advances.

1.3. Model Configuration Options

For post-combustion CO2 capture from flue gas, the amine-based CO_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_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.

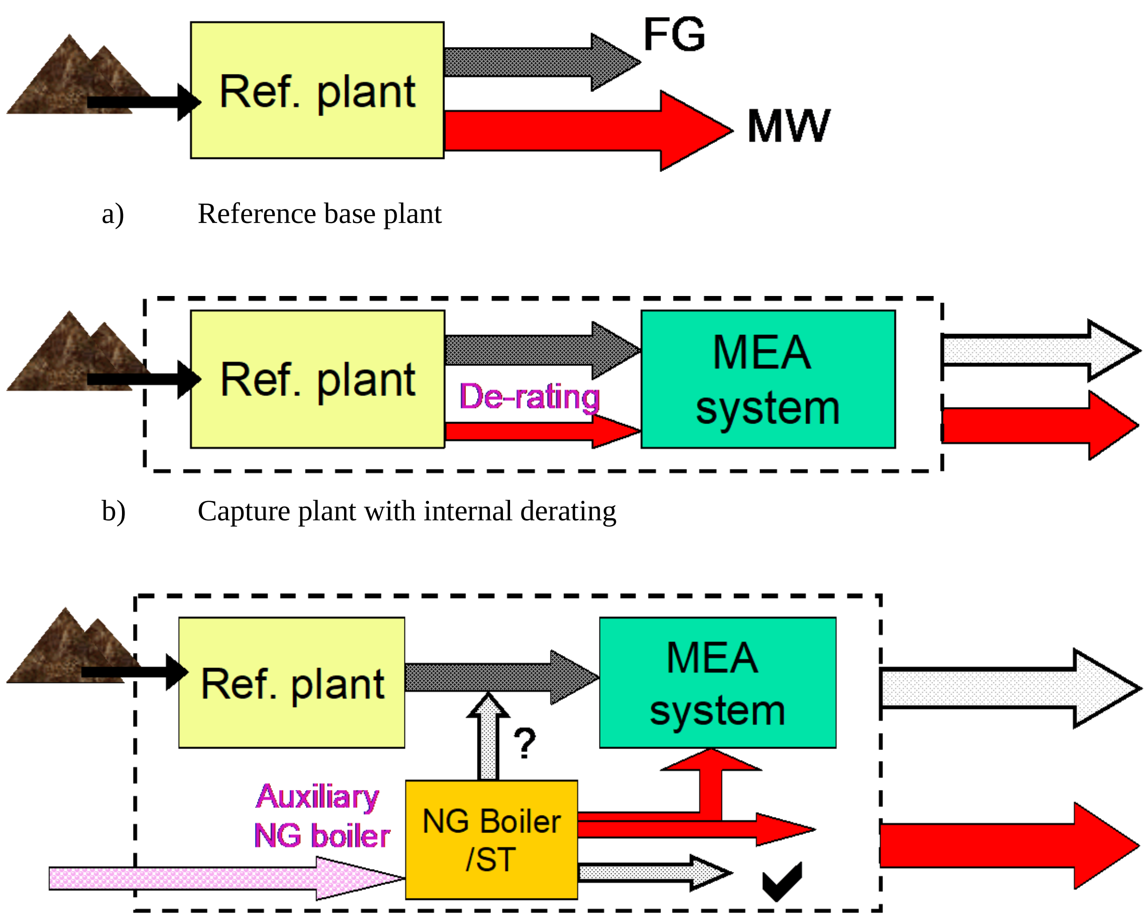
Auxiliary Natural Gas Boiler (w/ Steam Turbine): Another potential option to provide the energy for the amine system is by adding an auxiliary NG-fired boiler. Often it would be combined with a steam turbine which could generate some additional power (mainly used to supply electrical energy demand of CO_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_2 emissions (and NO_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_2 capture efficiency of the system is substantially lowered because of the additional CO_2 emissions from NG boiler. Also, the total NO_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_x removal before venting.

2. The CO_2 capture system maybe designed so as to capture CO_2 from the additional flue gas as well. In this case, the secondary flue gas stream (after cooling and NO_x removal, if required) maybe merged with main flue gas stream, before it enters the CO_2 capture system. If the NG fuel contains H_2S, the secondary flue gas may have to be treated for SO_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 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 CO_2 capture system treating the total flue gas. Thus, the 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 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. CO_2 Capture Plant Configuration Options



a) Reference base plant

b) Capture plant with internal derating

c) Capture plant with an auxiliary boiler and a secondary steam turbine

In terms of the configuration of the 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 ^oC. 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.

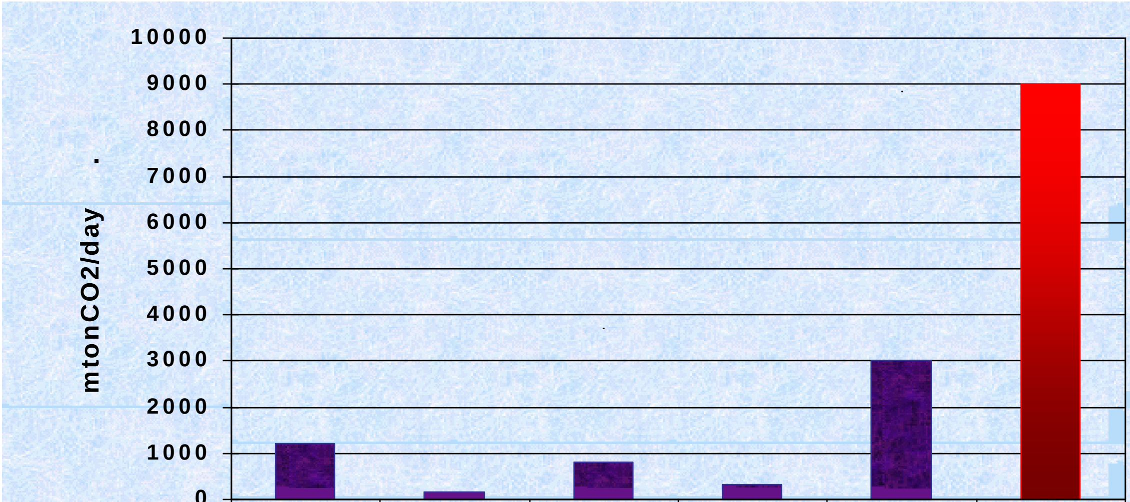
CO_2 transportation: The default mode of CO_2 transportation is via pipelines. The user can specify the distance over which CO_2 needs to be carried to, and the unit cost of CO_2 transportation. This module maybe expanded in future to include detailed parameters about pipeline transport and also other 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.



	EOR	Fertilizer	Soda Ash	Food-grade	Sleipner West Gasfield, Norway	500MW Coal-based Powerplant
mtonCO2/day	1200	150	800	320	3000	9000
Year	'80s	?	1978	?	1996	2010?
Process	Dow MEA	Dow MEA	Kerr-McGee	Dow MEA	MEA based	MEA based
Location	TX & NM	IGFC	NACC	NEEA, MA	Norway	?
Status	shutdown	operational	operational	operational	operational	prospective

Figure 4. Major Industrial Applications of CO_2 Capture Systems

All these plants capture CO_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_2S, CO_2) from natural gas streams (Herzog, 1999). The process was then adapted to treat flue gas streams for CO_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_2 capture. About 75%-95% CO_2 may be captured using this technology to yield a fairly pure (>99%) CO_2 product stream.

Today there are two main MEA-based processes available for commercial CO_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_2, or contain more than 1 ppm of H_2S, or contain less than 1% O_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_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 CO_2 from a gaseous stream. The system consists of two main elements, an absorber, where CO_2 is absorbed into a sorbent and a regenerator (or stripper), where CO_2 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 CO_2 in the flue gases to form a weakly bonded compound (carbamate). The scrubbed gases are then washed and vented to the atmosphere. The CO_2-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 CO_2 stream. The hot CO_2-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.

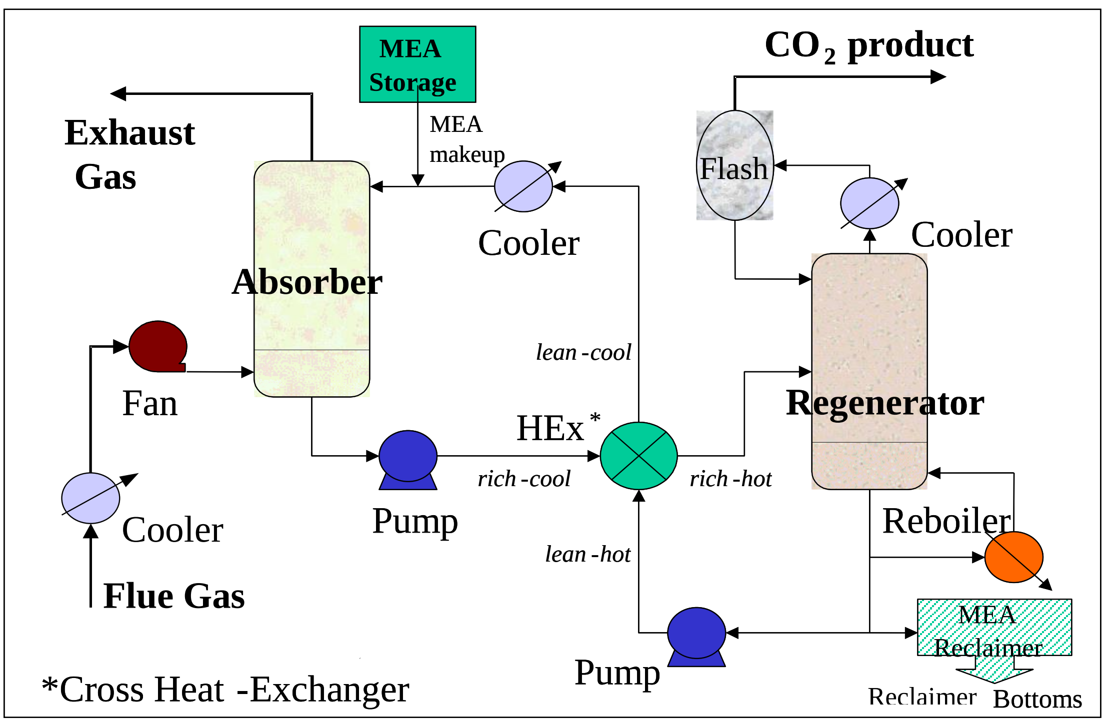


Figure 5. Flow Sheet for CO_2 Capture from Flue Gases using Amine-based System

The CO_2 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]

CO_2 Absorption: $2\ R\text{-}NH_2 + CO_2 \rightarrow R\text{-}NH_3^+ + R\text{-}NH\text{-}COO$

MEA Regeneration: $R\text{-}NH\text{-}COO^- + R\text{-}NH_3^+ + (\text{Heat}) \rightarrow CO_2 + 2\ R\text{-}NH_2$

Pure MEA (with R = $HO\text{-}CH_2CH_2$) is an "unhindered" amine that forms a weakly bonded intermediate called "carbamate" that is fairly stable. Only half a mole of CO_2 is absorbed per mole of amine, as shown in the CO_2 absorption equation above. On application of heat, this carbamate dissociates to give back CO_2 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 CO_2 per mole of amine, as shown in the CO_2 absorption equation below (Hezorg et al., 1997; Sartori, 1994).

CO_2 Absorption: $R\text{-}NH_2 + CO_2 + H_2O \rightarrow R\text{-}NH_3^+ + HCO_3$

MEA Regeneration: $HCO_3^- + R\text{-}NH_3^+ + (\text{less Heat}) \rightarrow CO_2 + H_2O + R\text{-}NH_2$

The regeneration of these amines requires lesser amount of heat energy as compared to the unhindered amines. But the CO_2 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 CO_2 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 SO_2 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 CO_2 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 SO_2 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 CO_2 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_2 transport. Recently, a 325-km pipeline carrying CO_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_2 disposal facility: Once the CO_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_2 to a storage site is typically assumed to be via pipeline. While the economic costs of CO_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_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_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_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_2-absorption capacity of the sorbent. Proprietary inhibitors are available that make the sorbent tolerant to oxygen. Flue gas NO_x is not a major problem since nitric oxide (NO) is the predominant form (~ 90-95%) of total NO_x in the flue gas, and does not react with inhibited amines (Suda et al., 1992; Leci, 1996). But, SO_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_2 (plus roughly 10-40 ppm NO_2). The interaction of SO_2 with CO_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 CO_2 removal options. Cost data also are relatively limited and often incomplete.

The rate of removal of CO_2 from flue gas using an amine scrubber depends on the gas-liquid mass transfer process. The chemical reactions that permit diffusion of CO_2 in the liquid film at the gas-liquid interface enhance the overall rate of mass transfer. So, the CO_2 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 CO_2 in an alkaline medium may be considered as a first order reaction. Higher CO_2 concentration thus improves the efficiency of the absorption system. Even at low concentration of CO_2, MEA has great affinity for CO_2. The solubility of CO_2 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 CO_2 capture and separation model is based on the *ProTreat* simulations while the CO_2 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 CO_2 capture and separation from flue gas

The CO_2 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 2. Protreat Parameter Ranges (total number of simulation runs: 1983)

No.	Parameter	Type	Units	Range
1	CO_2 content in flue gas (y_{CO2})	Input	mole %	3.5-13.5
2	Flue gas flow rate (G)	Input	kmole/hr	9000-24000
3	Inlet flue gas temperature (T_{fg})	Input	deg C	40-65
4	MEA concentration (C)	Input	wt %	15-40
5	Sorbent flow rate (L)	Input	kmole/hr	16000-70000
6	L/G	Input	-	0.73-5.56
7	Reboiler heat duty (Q)	Input	GJ/hr	95-664
8	Q/L	Input	MJ/kmole	2.4-22.5
9	CO_2 capture efficiency	Output	%	41.2-99.9
10	CO_2 product flow rate	Output	kmole/hr	333-2840
11	Lean sorbent CO_2 loading	Output	moleCO_2/mole MEA	0.05-0.34
12	Rich sorbent CO_2 loading	Output	moleCO_2/mole MEA	0.27-0.55
13	Absorber diameter	Output	ft	26-42
14	Regenerator diameter	Output	ft	12-42

15	Exhaust flue gas temperature	Output	deg C	40.4-71.6

The following set of parameters related to the design/configuration of the CO_2 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 CO_2 Compression

The concentrated CO_2 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^{o}C. The compressor efficiency, CO_2 product pressure and purity were used as the main control variables. These parameters were varied over the following ranges

- Compressor efficiency: 60-100 %
- CO_2 product pressure: 500-2500 psi
- CO_2 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 CO_2 capture system based on amine (MEA) scrubbing. This CO_2 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 CO_2 performance model:

- **Parameters from the "reference plant"**: These include the flow rate, temperature, pressure and composition of the flue gas inlet to the CO_2 absorber, and the gross power generation capacity of the power plant.
- **Parameters to configure the CO_2 system**: The CO_2 module provides a menu of options from which the user may select a CO_2 capture technology, CO_2 product pressure, mode and distance of CO_2 product transport, and CO_2 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 CO_2 system**: The main parameters include the CO_2 capture efficiency, MEA concentration, maximum and lean CO_2 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 CO_2 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 CO_2 system.

3.3.1. Parameters obtained from the "reference base plant"

The amine-based CO_2 capture system gets the following inputs from the (reference) base plant:

Gross plant size = MW_g

Net plant size (after env'l. controls) = MW_{noctl}

- 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 N_2, O_2, H_2O, CO_2, CO, HCl, SO_2, SO_3, NO, NO_2 and mass flow rate of particulates. The total molar flow rate of the flue gas is G, and the molar fraction of CO_2 in the flue gas is y_{CO2}.

Temperature of flue gas = T_{fg}

Plant capacity factor = PCF (%)

Annual hours of operation = HPY = (PCF/100)*365*24 hrs/yr

3.3.2. Parameters to configure the CO_2 system

These are the choices the user can make in order to configure the CO_2 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 CO_2 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 CO_2 system

Parameters controlling the performance of the CO_2 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.

CO_2 capture efficiency (η_{CO_2})

The overall CO_2 capture efficiency of the system is the fraction of CO_2 present in the incoming flue gas stream captured in this system.

$$\eta_{CO_2} = (\text{Moles } CO_2 \text{ in } - \text{ Moles } CO_2 \text{ out}) / (\text{Moles } CO_2 \text{ in})$$

Most of studies report the CO_2 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 CO_2 capture efficiency.

MEA concentration (C_{MEA})

The solvent used for CO_2 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 CO_2 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 CO_2. So, the regenerated (or lean) solvent contains some "left-over" CO_2. The level of lean solvent CO_2 loading mainly depends upon the initial CO_2 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 ration 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 CO_2 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 CO_2 capture system.

$$L = (L/G) \times (G)$$

Removal efficiency ($\eta_{acid\ gas}$) 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 CO_2 are given in Table 3.

Table 3. Removal Efficiency of Acid Gases Due to MEA Solvent (90% CO_2 removal)

Acid gas	removal efficiency (%)	MEA loss (mole MEA/mole acid gas)
SO_2	$\eta_{SO_2} = 99.5\%$	$n_{MEA,\ SO_2} = 2$
SO_3	$\eta_{SO_3} = 99.5\%$	$n_{MEA,\ SO_3} = 2$
NO_2	$\eta_{NO_2} = 25\%$	$n_{MEA,\ NO_2} = 2$
NO	$\eta_{NO} = 0$	$n_{MEA,\ NO} = 0$
HCl	$\eta_{HCl} = 95\%$	$n_{MEA,\ HCl} = 1$

Temperature of the flue gas entering the CO_2 capture system ($T_{fg,in}$)

The desirable temperature of the flue gas entering the CO_2 capture system is about 45-50 deg C. If a direct contact cooler is installed upstream of CO_2 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 CO_2 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 ($\dot{m}_{MEA,\ nom}$)

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:

$$\dot{m}_{MEA,\ polym} = = 50\% \text{ of } \dot{m}_{MEA,\ nom})$$

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

$$\dot{m}_{MEA,\ oxid} = 50\% \text{ of } \dot{m}_{MEA,\ nom}).$$

NH_3 Generation (n_{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).

$$\text{Rate of ammonia generation, } n_{NH_3} = 1 \ \frac{\text{mole } NH_3}{\text{mole MEA oxidized}}$$

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.]

$$n_{\text{MEA, organics}} = \left(1 \frac{\text{mole MEA}}{\text{mole org. acids}} \right)$$

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

Caustic (in the form of NaOH) is added in the reclaimer so that some of the MEA could be regenerated from HSS. $\dot{m}_{NaOH}$ 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.

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

$$= \text{no. of moles of caustic added}$$

$$= \dot{n}_{NaOH}$$

$$= \dot{m}_{NaOH} \,/\, (\text{Molecular Weight of NaOH})$$

$$= \dot{m}_{NaOH} \,/\, 40 \qquad (7)$$

Removal efficiency for particulates (η_{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 ($\rho_{sorbent}$)

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 ($\dot{m}_{act\text{-}C}$)

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 the suppliers). $\dot{m}_{act\text{-}C}$ 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.

$$n_{CO2} = (\eta_{CO_2} / 100)*(\text{Moles } CO_2 \text{ in}) = (\eta_{CO_2} / 100)*(G*y_{CO2})$$

Since the molecular weight of CO_2 is 44, the total amount of CO_2 captured (m_{CO2}, tonne/ hr) is

$$m_{CO2} = n_{CO2} * (44/1000)$$

CO_2 product purity (α)

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_{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 captued (m_{CO2}).

$$q_{regen} = Q / m_{CO2}$$

Enthalpy of regenerating steam (q_{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 maybe required. (h_{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%.

$$F_{HE} = 14\% \quad \text{i.e. } \mathrm{F_{HE}} = 0.14 \frac{\text{(kW.s) Electric}}{\text{(kJ) Heat}}$$

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 (ΔP_{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. (ΔP_{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 (η_{blower})

This is the efficiency of the fan/blower to convert electrical energy input into mechanical work output. Typically, the value of blower efficiency (η_{blower}) is about 75%.

Solvent head ($\Delta P_{solvent}$)

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. ($\Delta P_{solvent}$) 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 (η_{pump})

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

CO_2 product pressure (P_{CO_2})

The CO_2 product may have to be carried over long distances. Hence it is necessary to compress (and liquefy) it to very high pressures (P_{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_2 is about 1070 psig. The typically reported value of final pressure to which the product CO_2 stream has to be pressurized using compressors, before it is transported is about 2000 psig.

Energy required for CO_2 compression (e_{comp})

This is the electrical energy required (kWh per tonne CO_2) to compress a unit mass of CO_2 product stream to the designated pressure (P_{CO_2}) expressed in psig. Compression of CO_2 to high pressures takes lot of energy, and is a principle contributor to the overall energy penalty of a CO_2 capture unit in a power plant.

CO_2 compression efficiency (η_{comp})

This is the effective efficiency of the compressors used to compress CO_2 to the desirable pressure. Typically, the value of compressor efficiency (η_{comp}) is about 80%. It maybe noted that the energy requirement calculated from the performance equation (e_{comp}) has to be corrected by this efficiency factor in order to get the total energy required for CO_2 compression.

The following set of parameters are relevant only if the CO_2 capture system has been configured to include an auxiliary NG boiler to supply sorbent regeneration heat.

Heating value of natural gas (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 (ρ_{NG})

This is the density (lb/ft^3) of the natural gas used as fuel for the auxiliary boiler.

Average molecular weight of natural gas (mw_{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 (η_{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 (η_{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.

$$(L/G) = \exp(-1.4352 + 0.1239*y_{CO2} + 3.4863*\phi_{lean} + 0.0174*\eta_{CO2} - 0.0397*C + 0.0027*T_{fg,in})$$

[adj. $R^2 = 0.92$]

$$(Q/L) = \exp(2.5919 - 0.0059*y_{CO2} - 6.3536*\phi_{lean} + 0.0259*C - 0.0015*\eta_{CO2})$$

[adj. $R^2 = 0.96$]

$$(T_{fg,out}) = 41.15 + 0.062*T_{fg,in} + 1.307*y_{CO2} - 18.872*\phi_{lean} + 0.270*C$$

[adj. $R^2 = 0.92$]

$$(mw_{lean}) = 16.907 + 2.333*\phi_{lean} + 0.204*C$$

[adj. $R^2 = 0.95$]

$$(e_{comp}) = -51.632 + 19.207*\ln(P_{CO2} + 14.7)$$

[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_{CO2} = CO_2 concentration in the inlet flue gas (mole %)

ϕ_{lean} = lean sorbent CO_2 loading (mole CO_2/ mole MEA)

η_{CO2} = CO_2 capture efficiency (%)

C = MEA concentration in the sorbent (wt %)

$T_{fg,in}$ = Temperature of the flue gas entering the CO_2 absorber (deg C)

$T_{fg,out}$ = Temperature of the flue gas leaving the CO_2 absorber (deg C)

mw_{lean} = Average molecular weight of the lean sorbent (kg/ kmole sorbent)

e_{comp} = Unit energy requirement for CO_2 compression (kWh/ tonne CO_2)

P_{CO2} = Desired CO_2 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,

$$m_{CO2}\,(\text{tonne/hr}) = n_{CO2} \times (\text{MolWt})_{CO2}$$

$$= \eta_{CO2} \times n_{CO2,inlet} \times (\text{MolWt})_{CO2}$$

where,

$\dot{n}_{CO_2}$ = Total moles of CO2 captured (kmole CO_2/ hr)

$n_{CO2,inlet}$ = Molar flow rate of CO_2 in the inlet flue gas (kmole CO_2/hr)

$(\text{MolWt})_{CO2}$ = Molecular weight of CO_2 = 0.044 tonne/ kmole CO_2

- Net loss of MEA = MEA makeup requirement = $\dot{m}_{MEA,\ makeup}$

$$= \begin{pmatrix} \text{loss due to acid gas} \\ \textit{impurities} \end{pmatrix} + \begin{pmatrix} \text{loss due to} \\ \text{polymerization} \end{pmatrix} + \begin{pmatrix} \text{loss due to} & & \text{Gain in} \\ \text{HSS formation} & - & \text{Reclaimer} \end{pmatrix} + \begin{pmatrix} \text{loss with} \\ \text{fluegas exhaust} \end{pmatrix}$$

Estimation of total sorbent circulation flow rate: From the performance equations, we find

$$L/G = f\,(y_{CO2}, \phi_{lean}, \eta_{CO2}, C, T_{fg,in}), \text{ and } L = G*(L/G)$$

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

$$L_{tot,v} = \{G*(L/G)*mw_{lean} + \dot{m}_{MEA,\ makeup}*(100/C)\}*\rho_{sorbent}$$

- Waste generated from reclaimer:

$$m_{waste} = \begin{matrix} = \begin{pmatrix} \text{MEA lost due to} & + & \text{Total qty of} \\ \text{acid gases} & & \text{acid gases removed} \end{pmatrix} + \begin{pmatrix} \text{MEA loss due} \\ \text{to oxidation} \end{pmatrix} \\ + \begin{pmatrix} \text{MEA loss due to} \\ \text{HSS format}^{n} \\ -\text{Gain in reclaimer} \end{pmatrix} + \begin{pmatrix} \text{Caustic added} \\ \text{to} \\ \text{Reclaimer} \end{pmatrix} \end{matrix}$$

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

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

Typically, the reclaimer waste contains about 40% water.

- Activated carbon consumption

$$m_{act\text{-}C} = \dot{m}_{act\text{-}C} \times \dot{m}_{CO_2}\ \text{kg act-C/hr}$$

- Caustic Consumption in Reclaimer

$$m_{Caustic} = \dot{m}_{NaOH} \times \dot{m}_{CO_2} \quad \text{kg NaOH/ hr}$$

- Process Water requirement

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

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

$$(M_{pw}) = \dot{m}_{pw} \times MW_{net} \quad \text{tonne/hr}$$

- 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, SH_w = 4.2 kJ/kg °C

Specific heat of flue gas = SH_{fg} (Generally, this is around 1.2 kJ/kg °C)

Temperature rise in the cooling water (once through system) = ΔT_w

Drop in flue gas temperature = ΔT_{fg} = $(T_{fg,i} - T_{fg})$ °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} tonne/ hr

So, the required cooling water flow rate,

$$M_{cw} = m_{fg}*(\Delta T_{fg} / \Delta T_w)*(SH_{fg} / SH_w) \quad \text{tonne/hr}$$

Therefore, the total water requirement is:

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

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,

$$m_{steam} \text{ (tonne/hr)} = Q / q_{steam}$$

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 the following section 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_2 capture system is:

$$E_{CO_2,\,tot} = E_{regen} + E_{pumping} + E_{compr}$$

where,

E_{regen} = as explained in (9)

$$E_{pumping} = E_{blower} + E_{pump}$$

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

where Q_{fg} and ΔP_{fg} are expressed in ft^3/min and psi, respectively,

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

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

$$E_{compr} = e_{comp}*m_{CO2}/\eta_{comp}$$

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

Performance Parameter	Units	Data (Range)	Nominal Value	Unc. Representation (Distribution Function)
CO_2 removal efficiency	%	Mostly 90	90	-
SO_2 removal efficiency	%	Almost 100	99.5	Uniform(99,100)
NO_2 removal efficiency	%	20-30	25	Uniform(20,30)
HCl removal efficiency	%	90-95	95	Uniform(90,95)
Particulate removal eff.	%	50	50	Uniform(40,60)
MEA concentration	wt%	15-50	30	-

Lean solvent CO_2 loading	mol CO_2/mol MEA	0.15-0.30	0.22	Triangular(0.17,0.22,0.25)
Nominal MEA make-up	kg MEA/tonne CO_2	0.5-3.1	1.5	Triangular(0.5,1.5,3.1)
MEA loss (SO_2)	mol MEA/mol SO_2	2	2	-
MEA loss (NO_2)	mol MEA/mol NO_2	2	2	-
MEA loss (HCl)	mol MEA/mol HCl	1	1	-
MEA loss (exhaust gas)	ppm	1-4	2	Uniform (1,4)
NH_3 generation	mol NH_3/mol MEA oxidized	1	1	-
Caustic consumption in MEA reclaimer	kg NaOH/tonneCO_2	0.13	0.13	-
Activated carbon use	kg C/tonne CO_2	0.075	0.075	-
Cooling water makeup	m^3/tonne CO_2	0.5-1.8	0.8	Triangular (0.5,0.8,1.8)
Solvent pumping head	kPa	35-250	207	Triangular(150,207,250)
Pump efficiency	%	70-75	75	Uniform (70,75)
Gas-phase pressure drop	kPa	14-30	26	Triangular(14,26,30)
Fan efficiency	%	70-75	75	Uniform (70,75)
Equiv. elec. requirement	% regeneration heat	9-19	14[a]	Uniform (9,19)
CO_2 product purity	wt%	99-99.8	99.5	Uniform (99,99.8)
CO_2 product pressure	MPa	5.86-15.16	13.79	Triangular(5.86,13.79,15.16)
Compressor efficiency	%	75-85	80	Uniform (75,85)

[a] For retrofit applications, nominal value is 25.

4. Cost Model Development

The CO_2 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 CO_2 capture operation. Based on the same source, the maximum train size has been assumed to be 5000 tonnes per day of CO_2. Based on the actual CO2 capture rate ($\dot{m}_{CO_2}$) the minimum number of trains required to be installed (N_{min}) is determined. Different equipments have different maximum capacity limits. So, ($E_{n,i}$) defines the number of equipments required per train.

$E_{n,i}$: 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, CO_2 transport facility and CO_2 disposal facility, and they will be installed for the whole CO_2 capture unit. Hence "E_n" in this case, may be considered as ($1/N_t$) per train, where N_t is the total number of trains installed.

2. In case of CO_2 compressors, which have higher capacity (~ 7200 tpd CO_2), the number of compressors required is calculated accordingly. If N_c is the total number of CO_2 compressors installed, then the number of compressors installed per train may be stated as (N_c/N_t).

Different components of this system (Absorber, Regenerator, Flue gas blower etc.) are scaled, based on the flow ate 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 ate entering the CO_2 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

$$X = \frac{V}{E_{n,i} \bullet N_{\min}}$$

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

Each process area has a reference cost (C_{ref}) based on the source sited before, and the corresponding value of the scaling parameter (X_{ref}). The cost of the equipment is calculated using the reference values and the actual value of scaling parameter (X), based on the $6/10^{th}$ 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:

$C_{10,\,ref}$ = Cost of equipment (area 10)

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} = Minimum number of trains

$E_{n,10}$ = Number of equipment (10) per train

Minimum number of equipment installed, $Z_{10,min} = N_{min} \times E_{n,10}$

Total number of equipment installed, $Z_{10} = N_t \times E_{n,10}$

where,

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

So, the actual flue gas flow rate per train,

$X_{10} = Y / Z_{10,min}$

Therefore, the actual capital cost of absorber in this case may be estimated as

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

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.

$$C_{dcc} = C_{dcc,\,ref} \cdot \left(\frac{V_{fg}}{V_{fg,ref}} \cdot \frac{T_{fg}}{T_{fg,ref}} \right)^{0.6}$$

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

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

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.

$$C_{absorber} = C_{absorber,\,ref} \cdot \left(\frac{V_{fg,in}}{V_{fg,in,ref}} \cdot \frac{T_{fg,in}}{T_{fg,in,ref}} \right)^{0.6}$$

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.

$$C_{crossHEx} = \mathrm{C}_{\mathrm{crossHEx,\ ref}} \cdot \left(\frac{\mathrm{V}_{\mathrm{solvent}}}{V_{solvent,ref}} \right)^{0.6}$$

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

$$C_{regenerator} = \mathrm{C}_{\mathrm{regenerator,\ ref}} \cdot \left(\frac{\mathrm{V}_{\mathrm{solvent}}}{V_{solvent,ref}} \right)^{0.6}$$

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.

$$C_{reboiler} = \mathrm{C}_{\mathrm{reboiler,\ ref}} \cdot \left(\frac{\mathrm{V}_{\mathrm{solvent}}}{V_{solvent,ref}} \cdot \frac{\mathrm{M}_{\mathrm{steam}}}{M_{steam,ref}} \right)^{0.6}$$

It maybe noted that the ratio of mass flow rates of LP steam (M_{steam}/ $M_{steam,ref}$) 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.

$$C_{steam\_extractor} = \mathrm{C}_{\mathrm{steam\_extractor,\ ref}} \cdot \left(\frac{\mathrm{M}_{\mathrm{steam}}}{M_{steam,ref}} \right)^{0.6}$$

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_{NG_boiler} = \$15*(steam flow rate expressed in lb/hr)

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

C_{NG_boiler} = \$33000*(m_{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_{ST2} = \$300*(power generation expressed in kWe)

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

$$C_{ST2} = \$300000*(E_{ST2})$$

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.

$$C_{MEA\_reclaimer} = \mathrm{C}_{\mathrm{MEA\_reclaimer,\ ref}} \cdot \left(\frac{\mathrm{M}_{\mathrm{MEA\_makeup}}}{M_{MEA\_makeup,ref}} \right)^{0.6}$$

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

$$C_{solvent\_proc} = \mathrm{C}_{\mathrm{solvent\_proc,\ ref}} \cdot \left(\frac{\mathrm{V}_{\mathrm{solvent}}}{V_{solvent,ref}} \right)^{0.6}$$

CO_2 drying and compression unit: 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. Obviously, the size (and cost) of this unit will be a function of the CO_2 product flow rate, and maybe estimated as follows

$$C_{CO2\_compr} = \mathrm{C}_{\mathrm{CO2\_compr,\ ref}} \cdot \left(\frac{\mathrm{M}_{\mathrm{CO2}}}{M_{CO2,ref}} \right)^{0.6}$$

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 5. MEA Capital Cost Model Parameters and Nominal Values

	Capital Cost Elements	Value
A	Process Area Equipment Costs	$A_1, A_2, A_3, \ldots, A_{10}$
B	Total Process Facilities Cost (PFC)	ΣA_i
C	Engineering and Home Office	10% PFC
D	General Facilities	10% PFC
E	Project Contingency	15% PFC
F	Process Contingency	2% PFC
G	Total Plant Cost (TPC) = sum of above	B+C+D+E+F
H	Interest Costs During Constr.	Calculated
I	Royalty Fees	0.5% PFC
J	Pre-production (Fixed O&M)	1 month

K	Pre-production (Variable O&M Cost)	1 month
L	Inventory (startup) Cost	0.5% TPC
M	Total Capital Requirement (TCR)[a]	G+H+I+J+K+L

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 6. MEA O&M Cost Model Parameters and Nominal Values

O&M Cost Elements	Typical Value
Fixed O&M Costs	
Total Maintenance Cost	2.5% TPC
Maintenance Cost Allocated to Labor ($f_{maintlab}$)	40% of total maint. cost
Admin. & Support Labor Cost (f_{admin})	30% of total labor cost
Operating Labor (N_{labor})	2 jobs/shift
Variable O&M Costs	
Reagent (MEA) Cost	$1250/ mton
Water Cost	$0.8/ 1000 gallon
CO_2 Transport Cost	$0.04/ mton CO_2 km
CO_2 Storage/Disposal Cost	$5/ mton CO_2
Solid Waste Disposal Cost	$175/ mton waste

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

$$FOM = FOM_{labor} + FOM_{maint} + FOM_{admin}$$

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

$$FOM_{maint} = \Sigma_i (f_{maint})_i \times TPC_i \quad \text{where } i = \text{process area}$$

$$FOM_{admin} = f_{admin} \times (FOM_{labor} + f_{maintlab} \times FOM_{maint})$$

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

Cost of MEA reagent (VOM_{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.

$$VOM_{MEA} = M_{MEA,makeup} \times UC_{MEA} \times HPY$$

where, UC_{MEA} is the unit cost of MEA.

Cost of inhibitor ($VOM_{inhibitor}$): 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.

$$VOM_{inhibitor} = 0.2 \times VOM_{MEA}$$

Cost of other reagents ($VOM_{reagents}$): 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.

$$VOMreagents = VOM_{Caustic} + VOM_{act\text{-}C}$$

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

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

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

$$VOM_{waste} = M_{waste,total} \times UC_{waste} \times HPY$$

where, UC_{waste} is the unit cost of waste disposal for the reclaimer waste.

Cost of CO_2 transport ($VOM_{transport}$): Transportation of CO_2 product is assumed to take place via pipelines. The cost of CO_2 transport is estimated on the basis of two user specified parameters, viz., transportation distance (TD, in km) and unit cost of transport ($UC_{transport}$, \$/km mton CO_2), and CO_2 product flow rate (calculated result from performance model).

$$VOM_{transport} = M_{CO2} \times UC_{transport} \times TD \times HPY$$

Cost of CO_2 disposal ($VOM_{disposal}$): Depending upon the method of CO2 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_{disp}).

$$VOM_{disposal} = M_{CO2} \times UC_{disp} \times HPY$$

Cost of energy (VOM_{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_{noctl}) 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.

$$VOM_{energy} = E_{CO2,tot} \times HPY \times COE_{noctl}$$

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_{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_{water}) may vary depending upon the location of the power plant.

$$VOM_{water} = M_w \times UC_{water} \times HPY$$

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

$$VOM = VOM_{MEA} + VOM_{reagents} + VOMwaste + VOM_{transport} + VOM_{disposal} + VOM_{energy} + VOM_{water}$$

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).

$$\text{Cost of electricity, COE (\$/MWh)} = TRR / (MW_{net} * HPY)$$

Where,

TRR = Total annualized cost ($/yr)

MW_{net} = Net power generation capacity (MW)

HPY = 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:

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

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 CO2 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.

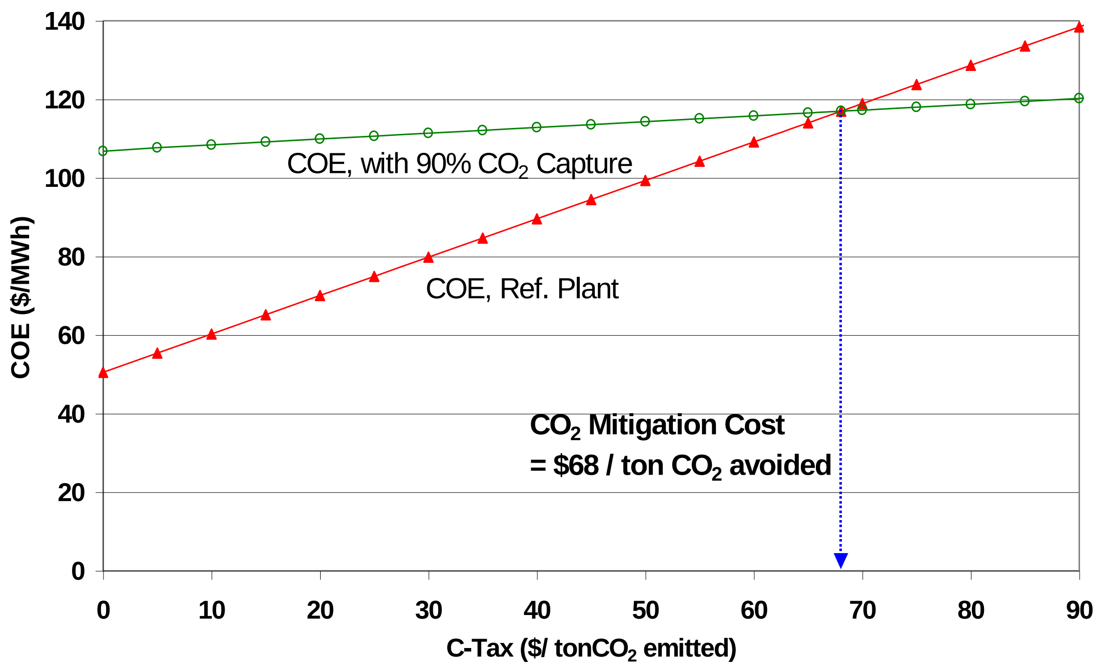


Figure 6. Cost of Electricity (COE) as A Function of Carbon Tax

5. Parameter and Model Updates for Advanced Amine-based CO_2 Capture System

5.1 PC Base Plant Updates

Here is the overview of base plant updates. The Base Plant module of the PC plant refers to plant with no environmental control systems. In the newest version of the IECM, a number of updates were made to the base plant. Specifically,

1. The default cooling system has been set to be a once-through system.
2. Ambient absolute air humidity is calculated in terms of given relative humidity.
3. A new process type, Steam Cycle has been added to the Set Parameter Base Plant screen.

4. The steam cycle heat rate has been updated based on the NETL 2007 Baseline Report and is now calculated as a function of boiler type, cooling technology, and CO_2 capture option.
5. Base plant power requirements are updated based on the NETL 2007 Baseline Report and are calculated as a function of boiler type and/or coal rank.
6. Leakage air at the preheater is updated based on recent guidelines (Babcock and Wilcox 2005).
7. The costs of all the pulverized coal power plant technologies were updated based on information available in the NETL 2007 Baseline Report.

The changes in the defaults are shown below in Table 7.

Table 7: Updated Base Plant Parameters

IECM Screen	Parameter	IECM 5.22 Defaults	IECM 6.2 Defaults	References
Configure Plant	Cooling System	Wet tower	Once-through	
Overall Plant	Ambient Air Humidity (lb H2O/lb dry air)	1.800e-02	9.879e-3 (Calc)	
Base Plant	Steam cycle heat rate (Btu/kWh)			
	Subcritical	7880	7790	NETL 2007
	Supercritical	7098	7359	NETL 2007
	Ultra-supercritical	6458	6705	
	Leakage Air at Preheater (%)	20	10	B&W 2005
	Base Plant Power Requirements (% of MWg)			NETL 2007
	Coal Pulverizer	0.6000	0.5105 (Calc)	
	Steam Cycle Pumps	0.6500	0.3100 (Calc)	
	Forced / Induced Draft Fans	1.500	1.408 (Calc)	
	Cooling System	1.800	0.4000 (Calc)	
	Miscellaneous	1.300	0.9900 (Calc)	

Here are the updated cost factors. For the new version of the IECM, the base process performance costs functions for all the pulverized coal power plant technologies were multiplied by a cost correction factor based on information available in the NETL 2007 Baseline Report (NETL 2007). This cost correction factor updates the costs of each technology. A summary table of the cost correction factors applied to each technology is shown in Table 8.

Table 8: Updated Cost Correction Factors by Technology

Technology	Cost Correction Factor
Base plant	1.12
SCR	0.99
TSP	0.86
FGD	1.59
Water System	1.01
CCS	1.60

5.2 Parameter and Model Updates for Advanced Amine-based CO_2 Capture System

In response to growing interest in large scale carbon dioxide (CO_2) capture, Fluor and Mitsubishi Heavy Industries (MHI) have developed commercially available advanced CO_2 capture systems based on solutions of aqueous amines (EPRI 2008). Fluor's most recent offering is the Econamine® FG+ process, which uses an aqueous mixture of monoethanolamine (MEA) and proprietary corrosion inhibitors (NETL 2007). The original IECM capture model is based on an MEA capture process that was representative of the state of the technology in 2002. In this updated version, a new CO_2 capture model based on the Fluor's Econamine® FG+ process is included in the IECM. For this new model, the performance of the original MEA model was adjusted to reflect process improvements in amine-based CO_2 capture. The corresponding costs have also been updated to reflect technology improvements and the current pricing environment.

MEA is costly to replace, has a significant regeneration heat requirement and can be corrosive. Improvements in MEA-based CO_2 capture processes therefore have been focused at lowering solvent losses, providing improved heat integration, and research into additives that inhibit corrosion allowing for the use of carbon steel instead of more expensive stainless steel (Roberts et al 2009). While many of these improvements are proprietary, the CO_2 capture system incorporated into the IECM can be modified to match the information publicly available on the performance and cost of these systems. The documentation that follows describes parameter updates made to the original MEA based CO_2 capture system that were applied for the advanced amine-based CO_2 capture system, model adjustments made for the advanced amine-based CO_2 capture system, menu updates for the capture models in the IECM in general, and a comparison between the MEA and advanced amine-based CO_2 capture system.

5.2.1 Parameter Updates for Advanced Amine-based CO_2 Capture System

The documentation in this section describes default parameter updates to the IECM MEA based CO_2 capture model that were made to represent the amine-based CO_2 capture processes in the DOE/NETL August 2007 Baseline Report, which is based on Fluor's Econamine® FG+ CO_2 capture system (NETL 2007). In the cases where the DOE/NETL August 2007 Baseline Report does not specify values for needed parameters, information was used from similar reports which included advanced amine-based CO_2 capture processes or were transferred from the original MEA based CO_2 capture model. The existing and updated cost and performance parameters are shown in each of the Tables below. The IECM 6.2 defaults are shown when a new supercritical PC plant is chosen that has been configured with a Hot-Side SCR, Cost-Side ESP, Wet FGD, and an Amine System with Advanced Amine (FG+) selected as the solvent.

Table 9: Set Parameters > CO_2 Capture > CO_2 Capture System Process > Config Menu

	Parameter	IECM 5.22 Default	IECM 6.2 Default	References
Config	Sorbent Used	MEA	Adv. Amine	
	Auxiliary Natural Gas Boiler	None	None	
	CO_2 Product Compressor Used?	Yes	Yes	
	Flue Gas Bypass Control	No Bypass	No Bypass	
	Direct Contact Cooler (DCC) Used?	Yes	Yes	
	SO_2 Polisher Used?	N/A	Yes	NETL 2007
	SO_2 Outlet Concentration (ppmv)	N/A	10	NETL 2007
	Temperature Exiting DDC (F)	122F	113F	IECM[1]

Table 10: Set Parameters > CO_2 Capture > CO_2 Capture System Process > Performance Menu

	Parameter	IECM 5.22 Default	IECM 6.2 Default	References
Performance	Maximum CO_2 Removal Efficiency (%)	90%	90%	
	Scrubber Removal Efficiency (%)	90%	90%	NETL 2007
	SO_2 Removal Efficiency (%)	99.50%	100%	NETL 2007
	SO_3 Removal Efficiency (%)	99.50%	99.50%	
	NO_2 Removal Efficiency (%)	25%	25%	
	HCl Removal Efficiency (%)	95%	95%	
	PM Removal Efficiency (%)	50%	50%	
	Max Train CO_2 Capacity (tons/hr)	230	230	
	# Absorbers	2	3	IECM[1]
	Spare Absorbers	0	0	
	Max CO_2 Compressor Capacity (tons/hr)	330	330	
	No.of Operating CO_2 Compressors	2	2	
	No.of Spare CO_2 Compressors	0	0	
	Scrubber Power Requirement (%MWg)	14.00%	9.187%	IECM[1]

Table 11: Set Parameters > CO_2 Capture > CO_2 Capture System Process > Capture Menu

	Parameter	IECM 5.22 Default	IECM 6.2 Default	References
Capture	Sorbent concentration (%)	30%	30%	NETL 2007
	Lean CO2 Loading (mol CO_2/mol sorb)	0.2	0.19	NETL 2007b
	Nominal Sorbent Loss (lb / ton CO_2)	3	0.6[2]	NETL 2007
	Liq/Gas Ratio (Ratio)	2.876	3.072	IECM[1]
	Ammonia Generation (mol NH_3 / mol sorb.)	1	1	
	Gas Phase Pressure Drop (psia)	2	1	NETL 2007b
	ID Fan Efficiency (%)	75%	75%	
	Makeup Water for Wash Section (% raw flue gas)	N/A	0.8	IECM[1]
	Regenerator Heat Requirement (Btu / lb CO2)	1975	1516	NETL 2007
	Steam Ht. Cont (Btu/lb Steam)	860.4	1397	NETL 2007
	Heat to Electricity Efficiency	14%	22%[3]	NETL 2007
	Solvent Pumping Head	30	30	
	Pump Efficiency	75%	75%	
	% Solids in Reclaimer Waste	40%	40%	
	Capture System Cooling Duty (ton H_2O/ton CO_2)	N/A	46.19	IECM[1]

Table 12: Set Parameters > CO_2 Capture > CO_2 Capture System Process > CO_2 Storage Menu

	Parameter	IECM 5.22 Default	IECM 6.2 Default	References
CO_2 Storage	CO_2 Product Pressure (psig)	2000	2000	
	CO_2 Compressor Efficiency (%)	80%	80%	
	CO_2 Unit Compression Energy (kWh/ton CO_2)	107	107	

Table 13: Set Parameters > CO_2 Capture > CO_2 Capture System Process > Retrofit Cost Menu

	Parameter	IECM 5.22 Default	IECM 6.2 Default	References
Retrofit Cost (All Units in retro $/new $)	SO2 Polisher/ Direct Contact Cooler	1	1	
	Flue Gas Blower	1	1	
	CO_2 Absorber Vessel	1	1	
	Heat Exchangers	1	1	
	Circulation Pumps	1	1	
	Sorbent Regenerator	1	1	
	Reboiler	1	1	
	Steam Extractor	1	1	
	Sorbent Reclaimer	1	1	
	Sorbent Processing	1	1	
	CO2 Drying and Compression Unit	1	1	
	Auxiliary Natural Gas Boiler	1	1	
	Auxiliary Steam Turbine	1	1	

Table 14: Set Parameters CO_2 Capture > CO_2 Capture System Process > Capital Cost Menu

	Parameter	IECM 5.22 Default	IECM 6.2 Default	References
Capital Cost	Construction Time (years)	3	3	
	General Facilities Capital (%PFC)	10	10	
	Engineering & Home Office Fees (%PFC)	7	7	
	Project Contingency Cost (%PFC)	15	15	
	Process Contingency Cost (%PFC)	5	5	
	Royalty Fees (%PFC)	0.5	0.5	
	Months of Fixed O&M (months)	1	1	
	Months of Variable O&M (months)	1	1	
	Misc. Capital Cost (%TPI)	2	2	
	Inventory Capital (%TPC)	0.5	0.5	
	TCR Recovery Factor (%)	100	100	

Table 15: Set Parameters > CO_2 Capture > CO_2 Capture System Process > O&M Cost Menu

	Parameter	IECM 5.22 Default	IECM 6.2 Default	References
O&M Cost	Sorbent Cost ($/ton)	1379	2142	NETL 2007
	Inhibitor Cost (% of MEA)	20	0	NETL 2007
	Activated Carbon Cost ($/ton)	1411	2000	NETL 2007
	Caustic (NaOH) Cost ($/ton)	666.6	413	NETL 2007
	Water Cost ($/kgal)	0.8874	1.03	NETL 2007
	Reclaimer Waste Disposal Cost ($/ton)	201.2	211.6	
	Electricity Price (Base Plant)	43.27	59.97	
	Number of Operating Jobs (jobs/shift)	2	2	
	Number of Operating Shifts (shifts/day)	4.75	4.75	
	Operating Labor Rate ($/hr)	24.82	33	NETL 2007

	Total Maintenance Cost (%TPC)	2.5	2.5	
	Maint. Cost Allocated to Labor (% total)	40	40	
	Administrative & Support Cost (% total labor)	30	30	
	CO2 Transportation Cost ($/ton)	2.4	2.38	IECM[1]
	CO2 Storage Cost ($/ton)	5.75	6.047	IECM[1]

[1]The IECM calculates updated values based on other values supplied as inputs. These values may change with different plant configurations.

[2]For a more thorough explanation and for the calculation of Nominal Sorbent Loss, see below.

[3] For the calculation of the updated Heat-to-Electricity Efficiency, see below.

5.2.2 Model Adjustments for Advanced Amine-based CO_2 Capture System

Updated Heat Integration Equation

In order to reflect the new regeneration heat requirement of the Advanced Amine CO_2 capture process, the original regression equation was adjusted by a scaling factor. In the original model, regeneration heat is calculated based on the following regression equation:

Regen Heat (Btu/ lb CO_2) = Sorbent Circulation (tons/hr)* exp(2.5919 + 0.0259 * Reagent Concentration (%) - 6.3536 * Lean CO_2 Loading (mol CO_2/sorb) - 0.0015 * Actual CO_2 Removal Efficiency (%) - 0.0059 * CO_2 Flue Gas (lb moles/hr)*100/Total Gas (lb moles/hr) / Sorbent Molecular Weight (lb/lb*mole)/ CO_2 Captured (tons/hr)* 429.9046

This regression equation was multiplied by a scaling factor of 0.7639 to approximately match the updated regeneration energy of 1516 Btu/ lb CO_2 currently available by advanced amine-based CO_2 capture systems, from 1984 Btu/ lb CO_2 available for conventional MEA systems.

Updated Heat-to-Electricity Conversion Efficiency Equation

In previous versions of the IECM, the Heat-to-Electricity Conversion Efficiency (or equivalence factor) was selected from a range of values in the literature as 14%. In this new version of the IECM, the Heat-to-Electricity Conversion Efficiency was estimated from data obtained using the NETL 2007 Baseline report (NETL 2007). In the NETL Baseline report, the following data is available for a subcritical plant without CO_2 Capture:

Case 9 – Subcritical Plant without CO_2 Capture,
Gross Plant Size: 583 MW
Coal Flow Rate: 437,699 lb coal/hr

Therefore, approximately 750 lb coal/hr is burned for each gross MW produced for this plant. For a subcritical plant with CO_2 Capture:

Case 10 – Subcritical Plan with CO_2 Capture,
Gross Plant Size: 680 MW
Coal Flow Rate: 646,589 lb coal/hr

Based on 750 lb coal/hr per gross MW produced and the 646,589 lb coal/hr flow rate of coal, the plant in Case 10 should produce approximately 862 MW gross absent the CO_2 capture system. Therefore, 862 MW-680MW = approximately 182 gross MW lost due to the CO_2 capture system

For NETL Case 10, the subcritical plant with the CO_2 capture system, 1,995,300 lb steam/hr of quality 1397.7 Btu/lb is diverted from the low pressure turbine and this steam flows to the reboiler of the advanced amine-based CO_2 capture unit. Therefore, the Heat-to-Electricity Conversion Efficiency can be calculated as:

$$\frac{182[\mathrm{MW}]}{1995300[\mathrm{lb\ steam/hr}]} * 1397.7[\mathrm{Btu/lb\ steam}] * 2.93\mathrm{E}^{-7}[\mathrm{MW/(Btu/hr)}] = 0.222 \text{ or } 22\%,$$

Where 2.93E-7 [MW/(Btu/hr)] equals the MW equivalent of 1 Btu/hr.

Similar results are shown using NETL Cases 11 and 12 for supercritical plants. In the new version of the IECM, the Heat-to-Electricity Conversion Efficiency as well as the steam quality parameters were updated for both the MEA and Advanced Amine CO_2 capture systems to reflect these results.

Updated Equivalent Electrical Loss Due to the Capture System

In previous versions of the IECM, the electrical equivalent loss (energy penalty) from the conventional CO_2 capture system was based on multiplying the regeneration energy requirement by the Heat-to-Electricity Conversion Efficiency. In the updated version of the IECM, this calculation was modified to reflect a more realistic process. The electrical equivalent loss (MW) due to the steam requirements in the reboiler are calculated in two steps, first by calculating the steam flow rate directly, and then by using this value and the Heat-to-Electricity Conversion Efficiency to calculate the electrical equivalent loss:

1. The steam flow rate is calculated as:

$$\dot{\mathrm{M}}_{\mathrm{Steam\ Flow}} = \frac{\mathrm{E}_{\mathrm{Regeneration}} * \mathrm{CO_2}_{\mathrm{Flow}}}{\mathrm{Enthalpy}_{\mathrm{Steam\ Inlet}} - \mathrm{Enthalpy}_{\mathrm{Steam\ Condensate}}}$$

Where:

$\dot{\mathrm{M}}_{\mathrm{Steam\ Flow}} = \mathrm{Steam\ Flow\ Rate\ [lb/hr]}$
$\mathrm{E}_{\mathrm{Regeneration}} = \mathrm{Regeneration\ Heat\ Energy\ [Btu/\ lb\ CO_2]}$
$\mathrm{CO_2}_{\mathrm{Flow}} = \mathrm{Flow\ Rate\ of\ CO_2\ Captured\ [lb\ CO_2/hr]}$
$\mathrm{Enthalpy}_{\mathrm{Steam\ Inlet}} = \mathrm{Enthaly\ of\ Steam\ at\ Reboiler\ Inlet\ [Btu/lb\ CO_2]}$
$\mathrm{Enthalpy}_{\mathrm{Steam\ Inlet}} = \mathrm{Enthalpy\ of\ Steam\ at\ Reboiler\ Outlet\ [Btu/lb\ CO_2]}$

2. The equivalent electrical loss (MW) is calculated as:

$$\mathrm{MW}_{\mathrm{Eq.}} = (\mathrm{Heat - to - Electricity\ Efficiency}) * \dot{\mathrm{M}}_{\mathrm{Steam\ Flow}} * \mathrm{Enthalpy}_{\mathrm{Steam\ Inlet}} * 2.97\mathrm{E}^{-7}$$

Where:

$MW_{Eq.}$ = Equivalent Steam Usage [MW]
Heat − to − Electricity Efficiency = Energy Conversion Efficiency [dimensionless]
$CO_{2_{Flow}}$ = Flow Rate of CO_2 Captured [lb CO_2/hr]
$Enthalpy_{Steam\ Inlet}$ = Enthaly of Steam at Reboiler Inlet [Btu/lb CO_2]
$2.97E^{-7}$ = Conversion Factor for Btu to MW [MW/[Btu/hr]]

Updated Base Plant Cost Equations for the Case with CO_2 Capture

In past versions of the IECM, the costs of the equipment in the Base Plant depended on the gross power produced (MW) by the power plant. In the updated version of the IECM for cases with CO_2 capture, the majority of equipment in the Base Plant is instead sized on the gross power produced (MW) plus the equivalent electrical loss (MW), which is called the Gross Power Produced Effective internally in the IECM. This new calculation accounts for the steam produced by the base plant that is not used to generate electricity but is instead used to regenerate amine in the CO_2 capture system. The only exception is the Turbine Island, which is still sized based the gross power produced. The Turbine Island does not need be sized based on the equivalent electrical loss because it is not part of the set of equipment that produces steam. This adjustment more realistically captures the cost equations of the base plant equipment for cases with CO_2 capture.

Updated Amine Usage for the Advanced Capture System

For the advanced amine-based CO_2 capture system, the Nominal Sorbent Loss (lb/ton CO_2) represents the amount of sorbent that has been lost due to unwanted polymerization and oxidation reactions. The total sorbent loss, which requires makeup sorbent (called Sorbent (lb/hr) on the Get Results, CO_2 Capture, Diagram screen), represents the Nominal Sorbent Loss (lb/ ton CO_2) minus the amount of CO_2 regenerated in the reclaimer.

5.2.3 Menu Updates to the IECM CO_2 Capture System

Updated CO_2 Capture, Config Menu

The IECM CO_2 Capture screens have been updated to include the changes listed in this document. The updated Advanced Amine process has been included in the Set Parameters, CO_2 Capture, Config Menu, and the user can now select between traditional MEA and Advanced Amine (FG+). An additional update to the Set Parameters, CO_2 Capture, Config Menu is the inclusion of a polishing unit intended to lower the SO_2 concentration in the feed gas. SO_2 can react with MEA to produce undesirable and irreversible byproducts, therefore causing a loss of amine in the system that requires makeup (NETL 2007). The polishing unit uses caustic to capture the SO_2, reducing the cost of makeup chemicals. The user can choose whether or not to include the polishing unit. The updated Set Parameters, CO_2 Capture, Config Menu is shown in Figure 7.

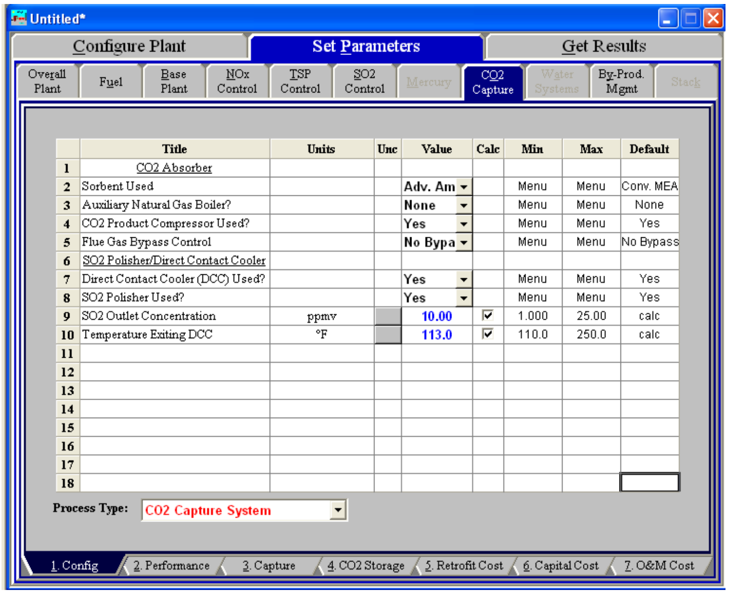


Figure 7. Updated CO2 Capture, Config Menu

Updated CO_2 Capture, Capture Menu

The CO_2 capture menu for the advanced amine system (Figure 8) has been updated to show only the Nominal Sorbent Loss. The Sorbent Oxidation Loss line has been collapsed into Nominal Sorbent Loss to reflect the lack of detailed information on this parameter available for advanced amine capture systems.

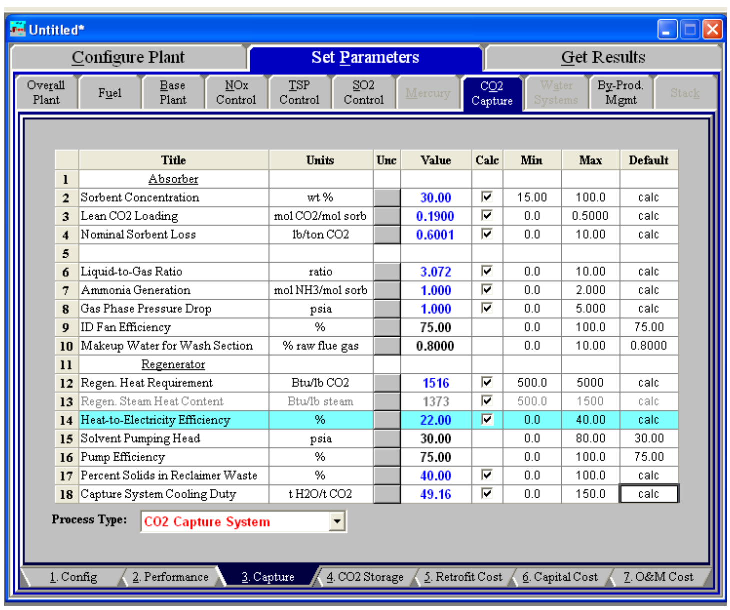


Figure 8. Updated CO2 Capture, Capture Menu

5.2.4 Case Studies of PC Plants

The NETL 2007 Baseline report presents four pulverized coal case studies: Subcritical PC plants with and without a CO_2 Capture System, carbon, and Supercritical PC plants with and without a CO_2 Capture System. For these case studies, parameters from the NETL 2007 Baseline Report (NETL 2007) were duplicated in the IECM, and the results from the updated IECM are presented.

Case Study #1: PC Supercritical without CO_2 Capture

For the Supercritical plant without a CO_2 Capture System, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table 16.

Table 16: IECM Parameters Changed for Case 1 - Supercritical Plants without CO_2 Capture

	Parameter	IECM Default	IECM Case 1
Configure Plant Overall Plant	NO_x Control	None	Hot-Side SCR
	Particulates	None	Fabric Filter
	SO_2 Control	None	Wet FGD
	Cooling System	None	Wet Cooling Tower
Fuel Properties	Fuel Name	Appalachian Medium Sulfur	Illinois #6
Overall Plant Performance	Capacity Factor (%)	75.00	85.00
	Ambient Air Temperature (Avg.) (°F)	77.00	70.00
Overall Plant Financing	Discount Rate (Before Taxes) (fraction)	0.103	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.148	0.164
	Plant or Project Book Life (years)	30.00	20.00
Overall Plant O&M Cost	Natural Gas Cost ($/mscf)	5.99	7.58
Base Plant Performance	Gross Electrical Output (MWg)	500	580.2
	Unit Type:	Sub-Critical	Supercritical
	Boiler Efficiency (%)	88.89	89.00
	Leakage Air at Preheater (% stoich.)	10	5.5
Base Plant Capital Cost	General Facilities Capital (%PFC)	10.00	2.41
	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	13.8
	Process Contingency Cost (%PFC)	0.3	0.0
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M Cost	Waste Disposal Cost ($/ton)	9.36	15.45
	Total Maintenance Cost (%TPC)	1.896	1.60
NO_x Control Performance	Actual NOx Removal Efficiency (%)	76.66	86.00
	Hot-Side SCR Power Requirement (% MWg)	0.5088	8.60E-3

NO_x Control Capital Cost	General Facilities Capital (%PFC)	10.00	2.41
	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	13.80
	Process Contingency Cost (%PFC)	0.3	0.0
	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO_x Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.10	99.90
	Fabric Filter Power Requirement (% MWg)	0.1958	1.70E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	2.41
	Engineering & Home Office Fees (%PFC)	5	9.37
	Project Contingency Cost (%PFC)	20	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.000	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.767	1.600
SO_2 Control Performance	Scrubber SO2 Removal Efficiency (%)	85.49	98.00
	Wet FGD Power Requirement (% MWg)	1.607	0.660
SO_2 Control Capital Cost	General Facilities Capital (%PFC)	10	2.41
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	2	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.3545	0.0
SO_2 Control O&M Cost	Total Maintenance Cost (%TPC)	4.306	1.60
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°F)	77.00	70.00
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	2.410
	Engineering & Home Office Fees (%PFC)	10	9.370
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.000	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0

Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

The key results of this case study are shown below in summary Table 20.

Case Study #2: PC Supercritical with CO_2 Capture

For the Supercritical plant with a CO_2 Capture System, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table 17.

Table 17: IECM Parameters Changed for Case 2 - Supercritical Plants with CO_2 Capture

	Parameter	IECM Default	IECM Case 2
Configure Plant Overall Plant	NO_x Control	None	Hot-Side SCR
	Particulates	None	Fabric Filter
	SO_2 Control	None	Wet FGD
	CO_2 Capture	None	Amine System
	Cooling System	None	Wet Cooling Tower
Fuel Properties	Fuel Name	Appalachian Medium Sulfur	Illinois #6
Overall Plant Performance	Capacity Factor (%)	75.00	85.00
	Ambient Air Temperature (Avg.) (°F)	77.00	70.00
Overall Plant Financing	Discount Rate (Before Taxes) (fraction)	0.103	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.148	0.175
	Plant or Project Book Life (years)	30.00	20.00
Overall Plant O&M Cost	Natural Gas Cost ($/mscf)	5.99	7.58
Base Plant Performance	Gross Electrical Output (MWg)	500	663.3
	Unit Type:	Sub-Critical	Supercritical
	Boiler Efficiency (%)	88.89	89.00
	Leakage Air at Preheater (% stoich.)	10	5.5
	Coal Pulverizer (% MWg)	0.5897	0.6700
Base Plant Capital Cost	General Facilities Capital (%PFC)	10.00	1.570
	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	16.38
	Process Contingency Cost (%PFC)	0.3	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M Cost	Waste Disposal Cost ($/ton)	9.36	15.45
	Total Maintenance Cost (%TPC)	1.896	1.60
NO_x Control Performance	Actual NOx Removal Efficiency (%)	76.66	86.00
	Hot-Side SCR Power Requirement (% MWg)	0.6294	1.100e-2

NO_x Control Capital Cost	General Facilities Capital (%PFC)	10.00	1.57
	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	10.00	16.38
	Process Contingency Cost (%PFC)	6.397	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO_x Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.10	99.90
	Fabric Filter Power Requirement (% MWg)	0.1958	1.70E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	1.57
	Engineering & Home Office Fees (%PFC)	5	9.37
	Project Contingency Cost (%PFC)	20	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.723	1.600
SO_2 Control Performance	Scrubber SO2 Removal Efficiency (%)	85.49	98.00
	Wet FGD Power Requirement (% MWg)	1.607	0.83
SO_2 Control Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	2	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.3545	0
SO_2 Control O&M Cost	Total Maintenance Cost (%TPC)	4.398	1.60
CO2 Capture Config	Sorbent Used	Conv. MEA	Adv. Amine (FG+)
CO_2 Capture Performance	Amine Scrubber Power Requirement (% MWg)	13.98	10.28
CO_2 Capture Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	7	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	5	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0

	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO_2 Capture O&M Cost	Reclaimer Waste Disposal Cost ($/ton)	211.6	0.0
	Total Maintenance Cost (%TPC)	2.5	1.600
	CO2 Transportation Cost ($/ton)	1.759	0.0
	CO2 Storage Cost ($/ton)	6.047	3.4
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°F)	77.00	70.00
	Power Requirement (% MWg)	2.8	2.800
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

The key results of this case study are shown below in summary Table 20.

Case Study #3: PC Subcritical without CO_2 Capture

For the Subcritical plant without a CO_2 Capture System, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table 18.

Table 18: IECM Parameters Changed for Case 3 - Supercritical Plants without CO_2 Capture

	Parameter	IECM Default	IECM Case 3
Configure Plant Overall Plant	NO_x Control	None	Hot-Side SCR
	Particulates	None	Fabric Filter
	SO_2 Control	None	Wet FGD
	Cooling System	None	Wet Cooling Tower
Fuel Properties	Fuel Name	Appalachian Medium Sulfur	Illinois #6
Overall Plant Performance	Capacity Factor (%)	75.00	85.00
	Ambient Air Temperature (Avg.) (°F)	77.00	70.00
Overall Plant Financing	Discount Rate (Before Taxes) (fraction)	0.103	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.148	0.164
	Plant or Project Book Life (years)	30.00	20.00
Overall Plant O&M Cost	Natural Gas Cost ($/mscf)	5.99	7.58
Base Plant Performance	Gross Electrical Output (MWg)	500	583.2
	Unit Type:	Sub-Critical	Sub-Critical

	Boiler Efficiency (%)	88.89	89.00
	Leakage Air at Preheater (% stoich.)	10	5.5
	Coal Pulverizer (% MWg)	0.6425	0.58
Base Plant Capital Cost	General Facilities Capital (%PFC)	10.00	2.515
	Engineering & Home Office Fees (%PFC)	6.50	9.3
	Project Contingency Cost (%PFC)	11.67	13.8
	Process Contingency Cost (%PFC)	0.3	0.0
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M Cost	Waste Disposal Cost ($/ton)	9.36	15.45
	Total Maintenance Cost (%TPC)	1.896	1.60
NO_x Control Performance	Actual NOx Removal Efficiency (%)	76.66	86.00
	Hot-Side SCR Power Requirement (% MWg)	0.5088	8.60E-3
NO_x Control Capital Cost	General Facilities Capital (%PFC)	10.00	2.515
	Engineering & Home Office Fees (%PFC)	6.50	9.3
	Project Contingency Cost (%PFC)	11.67	13.80
	Process Contingency Cost (%PFC)	0.3	0.0
	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO_x Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.10	99.90
	Fabric Filter Power Requirement (% MWg)	0.1958	1.70E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	2.515
	Engineering & Home Office Fees (%PFC)	5	9.30
	Project Contingency Cost (%PFC)	20	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.000	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.767	1.600
SO_2 Control Performance	Scrubber SO2 Removal Efficiency (%)	85.49	98.00
	Wet FGD Power Requirement (% MWg)	1.607	0.71
SO_2 Control Capital Cost	General Facilities Capital (%PFC)	10	2.515
	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	2	0.0
	Royalty Fees (%PFC)	0.5	0.0

	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.3545	0.0
SO_2 Control O&M Cost	Total Maintenance Cost (%TPC)	4.388	1.60
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°F)	77.00	70.00
	Power Requirement (% MWg)	1.41	1.410
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	2.515
	Engineering & Home Office Fees (%PFC)	10	9.30
	Project Contingency Cost (%PFC)	15	13.80
	Process Contingency Cost (%PFC)	0	0.0
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0.0
	Variable Operating Cost (months)	1.000	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	0.5	0.0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

The key results of this case study are shown below in summary Table 20.

Case Study #4: PC Subcritical with CO_2 Capture

For the Subcritical plant with a CO_2 Capture System, a number of default parameters in the IECM were changed so that these parameters matched the NETL 2007 Baseline Report. These changes are shown in Table 19.

Table 19: IECM Parameters Changed for Case 4 - Supercritical Plants with CO_2 Capture

	Parameter	IECM Default	IECM Case 4
Configure Plant Overall Plant	NO_x Control	None	Hot-Side SCR
	Particulates	None	Fabric Filter
	SO_2 Control	None	Wet FGD
	CO_2 Capture	None	Amine System
	Cooling System	None	Wet Cooling Tower
Fuel Properties	Fuel Name	Appalachian Medium Sulfur	Illinois #6
Overall Plant Performance	Capacity Factor (%)	75.00	85.00
	Ambient Air Temperature (Avg.) (°F)	77.00	70.00
Overall Plant Financing	Discount Rate (Before Taxes) (fraction)	0.103	1.0E-4
	Fixed Charge Factor (FCF) (fraction)	0.148	0.175
	Plant or Project Book Life (years)	30.00	20.00
Overall Plant O&M Cost	Natural Gas Cost ($/mscf)	5.99	7.58
Base Plant Performance	Gross Electrical Output (MWg)	500	681.3
	Unit Type:	Sub-Critical	Sub-Critical
	Boiler Efficiency (%)	88.89	89.00

	Leakage Air at Preheater (% stoich.)	10	5.5
	Coal Pulverizer (% MWg)	0.5897	0.7300
Base Plant Capital Cost	General Facilities Capital (%PFC)	10.00	1.570
	Engineering & Home Office Fees (%PFC)	6.50	9.37
	Project Contingency Cost (%PFC)	11.67	16.38
	Process Contingency Cost (%PFC)	0.3	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Fixed Operating Cost (months)	1.00	0.0
	Variable Operating Cost (months)	1.00	0.0
	Misc. Capital Cost (%TPI)	2.0	0.0
	Inventory Capital (%TPC)	6.00E-2	0.0
Base Plant O&M Cost	Waste Disposal Cost ($/ton)	9.36	15.45
	Total Maintenance Cost (%TPC)	1.896	1.60
NO_x Control Performance	Actual NOx Removal Efficiency (%)	76.66	86.00
	Hot-Side SCR Power Requirement (% MWg)	0.6294	1.100e-2
NO_x Control Capital Cost	General Facilities Capital (%PFC)	10.00	1.57
	Engineering & Home Office Fees (%PFC)	10.00	9.37
	Project Contingency Cost (%PFC)	10.00	16.38
	Process Contingency Cost (%PFC)	6.397	4.670
	Royalty Fees (%PFC)	7.00E-2	0.0
	Months of Fixed O&M (months)	1	0
	Months of Variable O&M (months)	1	0
	Misc. Capital Cost (%TPI)	2	0
	Inventory Capital (%TPC)	0.5	0
NO_x Control O&M Cost	Total Maintenance Cost (%TPC)	2.00	1.60
TSP Control Performance	Particulate Removal Efficiency (%)	99.10	99.90
	Fabric Filter Power Requirement (% MWg)	0.1958	1.40E-2
TSP Control Capital Cost	General Facilities Capital (%PFC)	1.00	1.57
	Engineering & Home Office Fees (%PFC)	5	9.37
	Project Contingency Cost (%PFC)	20	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
TSP Control O&M Cost	Total Maintenance Cost (%TPC)	1.723	1.600
SO_2 Control Performance	Scrubber SO2 Removal Efficiency (%)	85.49	98.00
	Wet FGD Power Requirement (% MWg)	1.607	0.89
SO_2 Control Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	2	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0

	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.3545	0
SO_2 Control O&M Cost	Total Maintenance Cost (%TPC)	4.398	1.60
CO_2 Capture Config	Sorbent Used	Conv. MEA	Adv. Amine (FG+)
CO_2 Capture Performance	Amine Scrubber Power Requirement (% MWg)	13.98	11.04
CO_2 Capture Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	7	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	5	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.00	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
CO_2 Capture O&M Cost	Reclaimer Waste Disposal Cost ($/ton)	211.6	0.0
	Total Maintenance Cost (%TPC)	2.5	1.600
	CO2 Transportation Cost ($/ton)	1.759	0.0
	CO2 Storage Cost ($/ton)	6.047	3.4
Water Systems Performance	Ambient Air Temp (Dry Bulb Avg.) (°F)	77.00	70.00
	Power Requirement (% MWg)	2.8	3.14
Water Systems Capital Cost	General Facilities Capital (%PFC)	10	1.57
	Engineering & Home Office Fees (%PFC)	10	9.37
	Project Contingency Cost (%PFC)	15	16.38
	Process Contingency Cost (%PFC)	0	4.670
	Royalty Fees (%PFC)	0.5	0.0
	Fixed Operating Cost (months)	1.0	0
	Variable Operating Cost (months)	1.000	0
	Misc. Capital Cost (%TPI)	2.0	0
	Inventory Capital (%TPC)	0.5	0
Water Systems O&M Cost	Total Maintenance Cost (%TPC)	2.0	1.600

The key results of this case study are shown below in summary Table 20.

Summary Results from Case Studies 1-4

The key results from Case Studies 1-4 are shown below in Table 14. The performance parameters in terms of net plant efficiency for the same gross power plant size match closely, as do the capital costs on a $/kW-net basis. The revenue required for the IECM for each of these cases is somewhat lower than in the NETL 2007 Baseline Report, owing primarily to an escalation in coal prices for all four cases in the Baseline Report, which is not done in this analysis in the IECM.

Table 20: Summary of Case Study Results

Cases	Gross Output (MW)		Net Plant Efficiency (%)		Capital Cost ($/kW -net)		Revenue Required ($/MWh)	
	IECM	NETL	IECM	NETL	IECM	NETL	IECM	NETL
PC Supercritical	580.2	580.2	39.1%	39.1%	1601	1575	60.4	63.3
PC Supercritical + CCS	663.3	663.4	27.1%	27.2%	2857	2870	108.4	114.8
PC Subcritical	583.2	583.3	36.8%	36.8%	1541	1549	60.1	64.0
PC Subcritical + CCS	681.3	679.9	24.8%	24.9%	2935	2895	112.5	118.8

5.2.5 Cost Sensitivity

A cost sensitivity analysis was done for different steam cycles (Subcritical, Supercritical, and Ultra Supercritical), for different coal types (Appalachian Medium Sulfur, Illinois #6, and Wyoming Powder River Basin), and for plants with and without CO_2 capture across a range of net plant sizes (250MW to 750MW). In each of these sensitivity studies, the defaults for the IECM were used with an SCR, ESP, and FGD system, and with an Advanced Amine CO_2 Capture System for the case with CO_2 capture. The results are presented below.

Cost Sensitivity to Steam Cycle Type

In the sensitivity analysis for the steam cycle, an IECM default plant was built with an SCR, ESP, and FGD system, but without a CO_2 capture system, and the coal used was Illinois #6. The net plant size was varied between 250 and 750 MW. The plant efficiencies were approximately constant through this range of plant sizes, with the Subcritical plant having an efficiency of 36.4%, the Supercritical plant having an efficiency of 38.7%, and the Ultra Supercritical plant having an efficiency of 42.7% (all HHV). The results from the sensitivity studies for Capital Cost and Revenue Required are shown in Figure 9 and Figure 10 below.

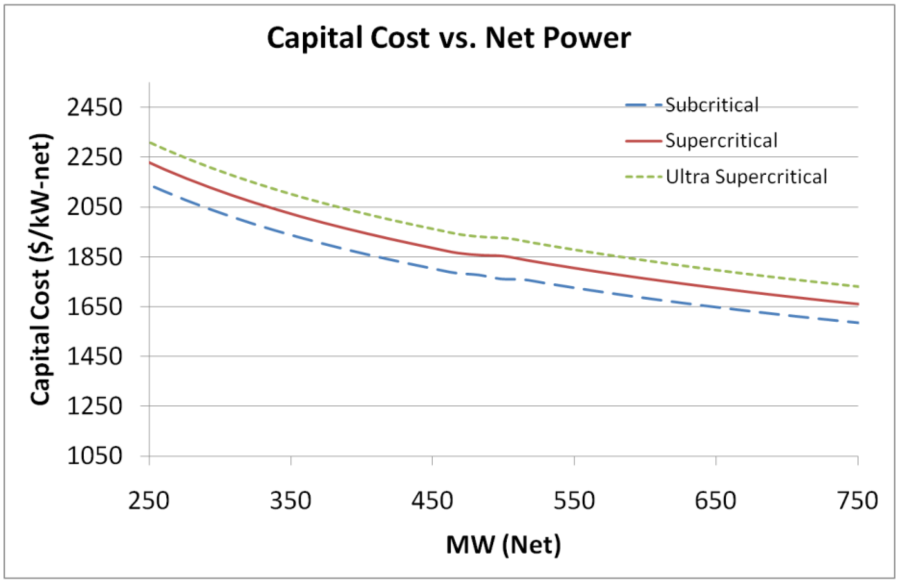


Figure 9. Capital Costs by Steam Cycle Type vs. Net Power Output

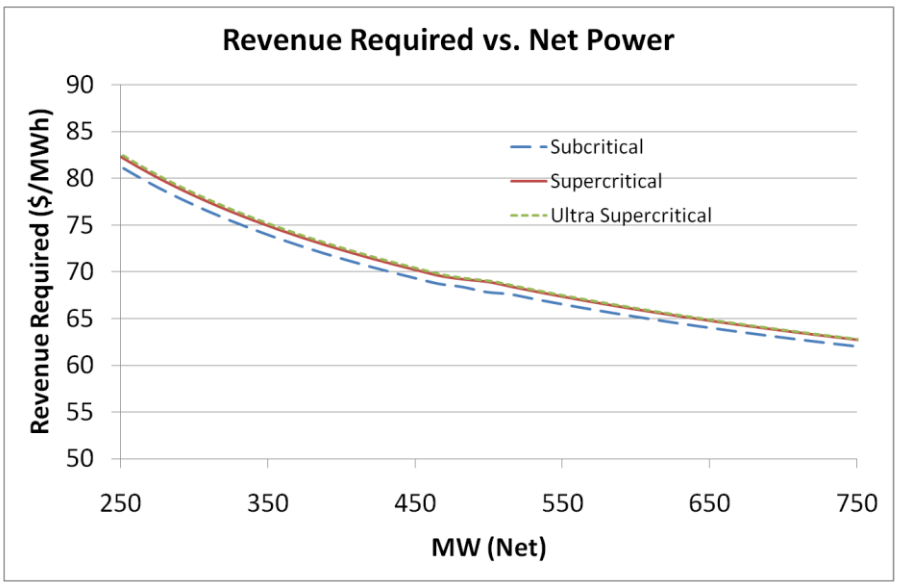


Figure 10. Revenue Required by Steam Cycle Type vs. Net Power Output

Sensitivity to Coal Type

In the sensitivity analysis for coal type, an IECM default plant was built with an SCR, ESP, and FGD system, but without a CO_2 capture system, and a Supercritical steam cycle was used. Three coals were chosen for this analysis, Appalachian Medium Sulfur, Illinois #6, and Wyoming Powder River Basin. The net plant size was varied between 250 and 750 MW. The results from the sensitivity studies for Capital Cost and Revenue Required are shown in Figure 11 and Figure 12 below.

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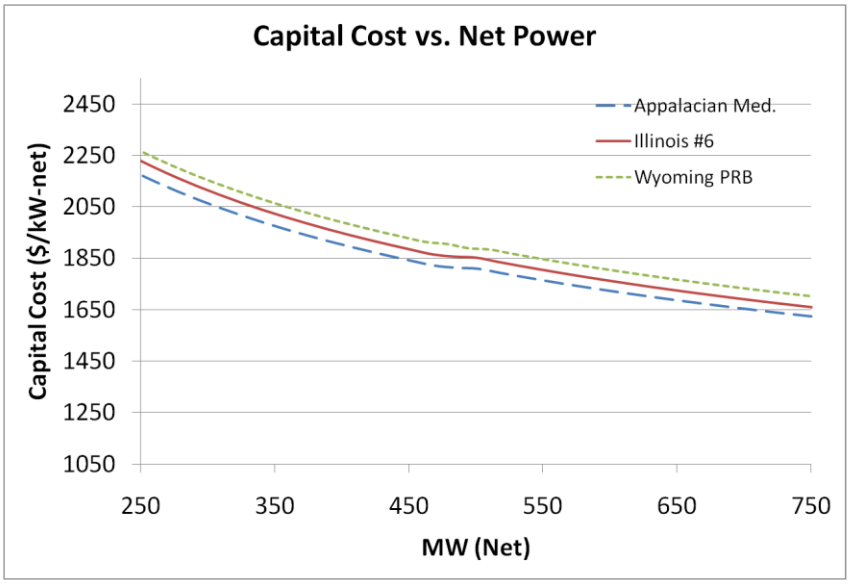

Figure 11. Capital Costs by Coal Type vs. Net Power Output

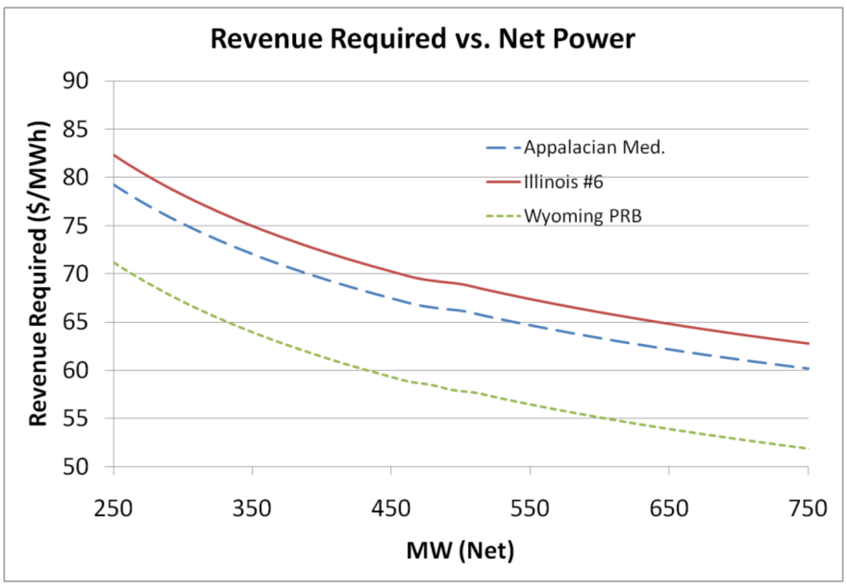

Figure 12. Revenue Required by Coal Type vs. Net Power Output

Sensitivity to CO_2 Capture Technology

In the sensitivity analysis for CO_2 capture technology, an IECM default plant was built with an SCR, ESP, and FGD system, and a Supercritical steam cycle was used. The net plant size was varied between 250 and 750 MW with and without a CO_2 capture system. The results from the sensitivity studies for Capital Cost and Revenue Required are shown in Figure 13 and Figure 14 below. The costs for plants with CO_2 Capture generally decrease with increasing plant size until a new train is required in the CO_2 capture system, at which point the cost rises slightly.

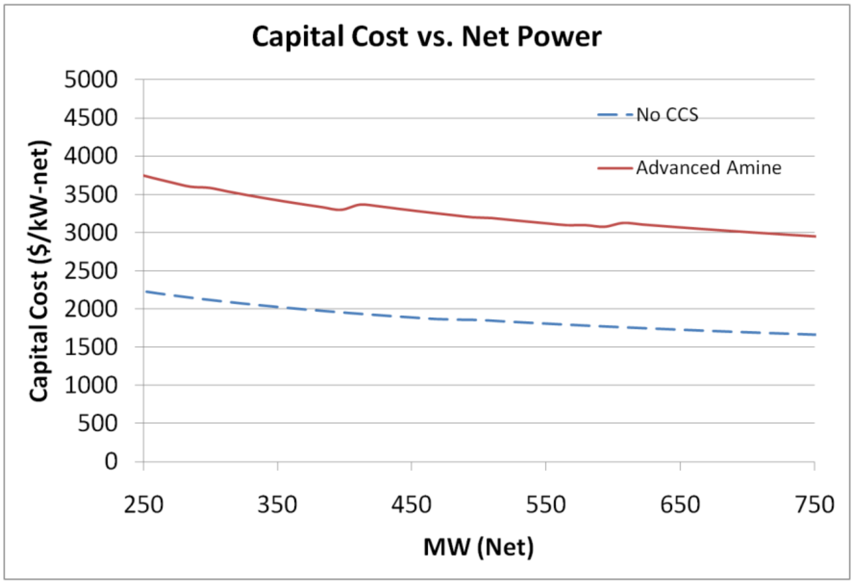

Figure 13. Capital Cost vs. Net Power Output with and without CCS

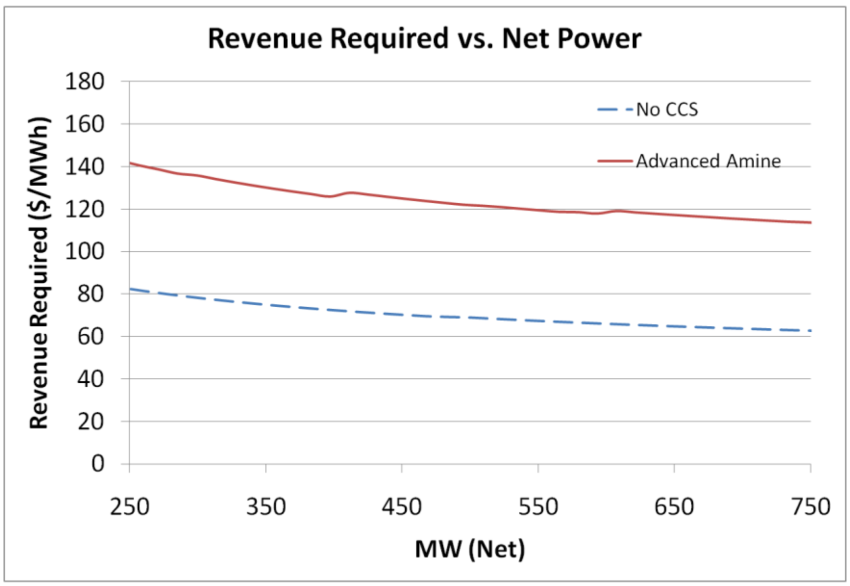

Figure 14. Revenue Required vs. Net Power Output with and without CCS

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