JV TASK 106 – FEASIBILITY OF CO_2 CAPTURE TECHNOLOGIES FOR EXISTING NORTH DAKOTA LIGNITE-FIRED PULVERIZED COAL BOILERS

Final Report

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1.0 INTRODUCTION

1.1 Project Description and Objectives

The goal of this project is to provide a technical review and evaluation of various carbon dioxide (CO_2) capture technologies, with a focus on the applicability to lignite-fired facilities within North Dakota. The motivation for the project came from the Lignite Energy Council's (LEC's) need to identify the feasibility of CO_2 capture technologies for existing North Dakota lignite-fired, pulverized coal (pc) power plants. A literature review was completed to determine the commercially available technologies as well as to identify emerging CO_2 capture technologies that are currently in the research or demonstration phase. The literature review revealed few commercially available technologies for a coal-fired power plant. CO_2 separation and capture using amine scrubbing have been performed for several years in industry and could be applied to an existing pc-fired power plant. Other promising technologies do exist, but many are still in the research and demonstration phases. Oxyfuel combustion, a technology that has been used in industry for several years to increase boiler efficiency, is in the process of being tailored for CO_2 separation and capture.

These two technologies were chosen for evaluation for CO_2 separation and capture from coal-fired power plants. Although oxyfuel combustion is still in the pilot-scale demonstration phase, it was chosen to be evaluated at LEC's request because it is one of the most promising emerging technologies. As part of the evaluation of the two chosen technologies, a conceptual design, a mass and energy balance, and an economic evaluation were completed.

1.2 Background

1.2.1 CO_2 Is an Environmental Concern

In 1992, international concern about climate change led to the United Nations Framework Convention on Climate Change (UNFCCC). The ultimate objective of that convention was the "stabilization of greenhouse gas concentrations in the atmosphere at a level that mitigates anthropogenic interference with the climate system" (1). Research by the U.S. Department of Energy (DOE) and the International Energy Agency (IEA) has suggested that carbon separation and sequestration can play an important role in reducing CO_2 in the atmosphere in the first part of the twenty-first century (2).

Currently, global warming is perceived by many as the largest environmental challenge facing the world. An increased level of CO_2 in the atmosphere has been interpreted as the dominant contributor to the apparent increase in global warming. The primary sources of anthropogenic CO_2 are fossil-fueled power plants, automobile engines, and furnaces used in residential and commercial buildings. Ninety-seven percent of anthropogenic CO_2 emissions come from energy-related activities (3). CO_2 emissions from coal-fired power plants contributed more than one-third of the anthropogenic CO_2 emissions in the United States in 2004. A breakdown of the stationary U.S. CO_2 emissions is outlined in Table 1, which shows that CO_2 from coal-fired electricity-producing utilities is the single largest contributor of all stationary emitters. Because of the abundant supply of coal, especially lignite and subbituminous and bituminous coals, the United States will rely on the use of fossil fuels for its energy needs for many years to come, thus sustaining or increasing the level of CO_2 emissions. Since lignites produce more CO_2 per unit of energy compared to the other ranks of coal, they will be the most impacted by any move to force CO_2 emission reductions from power plants.

1.2.2 CO_2 Capture

The three main options for reducing CO_2 emissions from fossil fuel-based energy systems are 1) increasing the fuel conversion efficiency, 2) switching to a fuel with a lower fossil carbon content, and 3) capturing and storing the CO_2 emitted from the fossil fuel (4). Options 1 and 2 are currently not feasible options as the United States relies, and will continue to rely, heavily on coal for energy production. Reduction of anthropogenic CO_2 emissions is focused on CO_2 separation and subsequent sequestration, which includes capture and separation, transportation, and storage. Sixty percent of the total cost for CO_2 sequestration occurs in the capture and separation step (2). It is technically feasible to separate CO_2, but the costs associated with the method are too high to be practical.

Table 1. Annual U.S. CO_2 Emissions

Sources	U.S. Total Tonnes
Power Generation[1]	2,239,700,000
Coal[1]	1,868,400,000
Natural Gas[1]	299,100,000
Oil[1]	72,200,000
Industries	324,789,000
Refinery[2]	184,918,000
Iron and Steel[3]	54,411,000
Cement[3]	42,898,000
Ammonia[3]	17,652,000
Aluminum[3]	4,223,000
Lime[3]	12,304,000
Ethanol[3]	8,383,000
Total	2,564,489,000

[1] U.S. Environmental Protection Agency Greenhouse Gas Inventory Sector Analysis (3).

[2] Estimate from U.S. DOE, 2002 BPD totals (5).

[3] U.S. EPA (2004) Greenhouse Gas Inventory Industrial Process Analysis (6).

1.2.3 CO_2 Regulations in the United States

Vast numbers of state and federal regulations in the United States deal with emissions from industrial and energy generation facilities. To date, none of these regulations has classified CO_2

as a pollutant, and no regulations currently govern CO_2 emissions into the atmosphere. The United States has not yet promulgated any regulations addressing CO_2 emissions. However, its Global Climate Change Initiative has set a goal to reduce greenhouse gas intensity (defined as the CO_2 emissions divided by the gross domestic product) by 18% by 2012 through the support of voluntary efforts by industry.

Only the inventory list for the Toxic Substances Control Act (TSCA) of 1976, the National Institute of Occupational Safety and Health (NIOSH) confined space hazard classification system, and the Federal Emergency Management Agency (FEMA) hazardous materials guide treat CO_2 as a hazardous substance to the extent that any concentrated, pressurized, or cryogenic gas poses a danger. Surface risks of CO_2 exposure are typically handled by state environmental health and safety regulatory agencies. For human health, the Occupational Safety and Health Administration (OSHA) has specified the maximum average exposure of CO_2 over an 8-hour workday at 0.5 wt% (5000 ppm). Most industrial and safety regulations for CO_2 focus on engineering controls and specifications for transportation, storage containers, and pipelines (7).

2.0 CO_2 CAPTURE TECHNOLOGIES

There are three opportunities for capture of CO_2 from a fossil fuel combustion system: before, during (through combustion modification), and after combustion.

2.1 Precombustion Removal

Precombustion removal refers to near-complete capture of CO_2 prior to fuel combustion and is usually implemented in conjunction with gasification (of coal, coke, waste, residual oil) or steam/partial oxidation reforming of natural gas to produce syngas. Syngas contains CO and H_2. Subsequent conversion via the water–gas shift reaction produces CO_2 from CO, resulting in H_2-rich syngas. This syngas (often with N_2 added for temperature control) can be combusted in gas turbines, boilers, or furnaces. Figure 1 is a flow sheet showing precombustion CO_2 removal.

Typical CO_2 stream concentrations before capture are 25 to 40 vol% at pressures of 363 to 725 psia. The high partial pressure of CO_2, relative to that in combustion flue gas, enables easier separation through solvent scrubbing. In refineries and ammonia production facilities, where H_2-rich syngas is produced by gas reforming, CO_2 is recovered during acid gas removal using chemical solvents (e.g., Benfield or MDEA [methyldiethanolamine] processes). Pressure swing adsorption (PSA) is also used, but the CO_2-rich stream may have significant residual fuel value that makes it attractive for in-plant use.

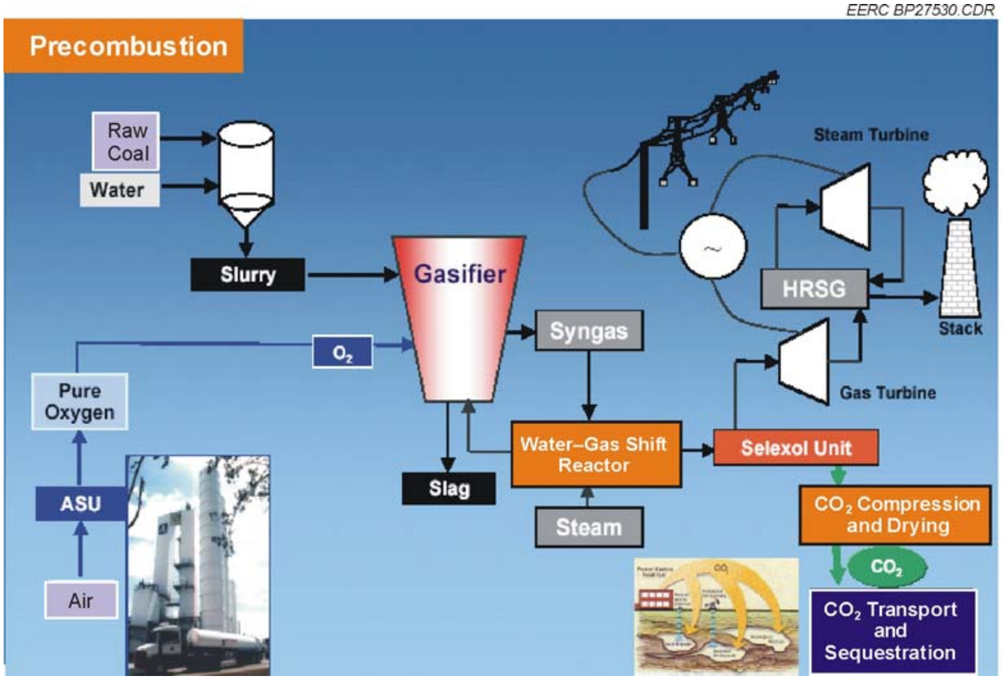

Figure 1. Schematic of precombustion CO_2 capture (8).

2.2 Removal During Combustion

Substitution of oxygen for all of the combustion air has been proposed to produce a CO_2-rich flue gas requiring minimum separation for use or sequestration. Conventional air combustion processes in boilers or gas turbines produce flue gas that contains predominantly N_2 (>80 vol%) and excess O_2 in addition to CO_2 and water. Separation technologies must separate CO_2 from these other components. If the air is replaced by oxygen, the nitrogen content of the flue gas approaches zero (assuming minimal air leakage into the system), and the flue gas contains predominantly CO_2 along with a small amount of excess oxygen and combustion water. The CO_2 can be recovered by compressing and cooling, followed by dehydration. The adiabatic flame temperature can be moderated by recirculating a part of the recovered CO_2. This is shown in Figure 2.

The levels of noncondensable impurities and thermodynamics limit recovery of CO_2 and affect the purity of the product stream. The concentration of CO_2 can be targeted to a specific intended end-use application such as sequestration. For enhanced coalbed methane recovery (ECBM) or saline aquifer sequestration, only condensation of moisture may be required because some constituents (e.g., N_2) can be present and a supercritical, dense-phase fluid is not required. Under this scenario, zero emissions would be possible. Where a supercritical fluid is required for EOR or Deep Reservoir injection, noncondensable contaminants such as N_2, NO_x, O_2, and Ar are removed by flashing in a gas–liquid separator.

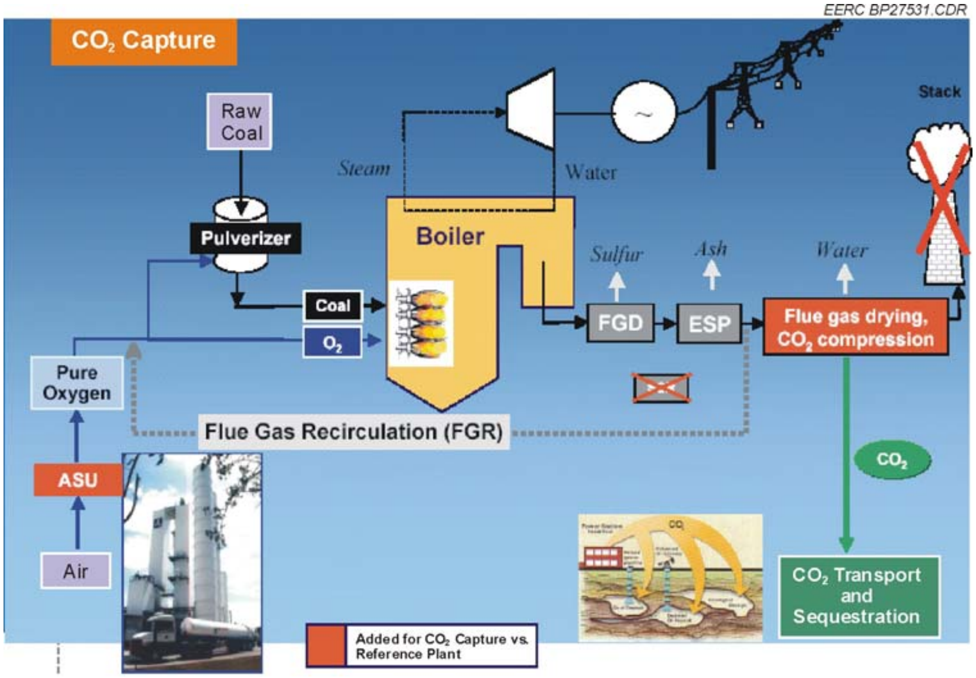

Figure 2. Schematic of an oxygen combustion system (8).

There are several advantages to oxygen combustion. The volume of flue gas reaching downstream systems is one-third to one-fifth that of conventional coal boilers. The process produces a flue gas stream containing more than 80 vol% CO_2, depending upon the fuel composition, purity of oxygen from air separation, and air leakage into the boiler. Impurities such as SO_2, NO_x, particulate, and mercury become concentrated in the flue gas, thus reducing capital and operating costs for contaminant removal. NO_x may be low enough to eliminate further control, and capital and operating cost savings (for control systems) may offset air separation capital and operating costs.

Issues with oxygen combustion center principally around the high cost for air separation, which is currently attainable at a very large scale only by cryogenic distillation. Relative to coal gasification, combustion requires up to three times the amount of oxygen because all of the carbon is converted to CO_2. The air separation unit (ASU) capacity (and parasitic power load) likewise will be commensurately larger. Other issues include expected lower flue gas exit temperature (that may increase the risk of low-temperature corrosion from condensation of sulfuric acid), burner operation, flame stability, levels of unburned carbon, flame luminosity and length, and changes in slagging/fouling characteristics under the different atmosphere.

Retrofit applications would be designed to maintain the same steam outlet conditions. The higher heat capacity of the gas should potentially facilitate greater heat absorption while producing lower flue gas temperature. Higher heat absorption would result in higher boiler efficiency, but this would be offset by higher auxiliary power load for fan power to the recycle gas for temperature control.

Development efforts involving conventional pulverized coal testing with oxygen combustion are at the scale of several hundred kilowatts and less. Developers and testing organizations include CANMET, Mitsui Babcock, American Air Liquide, Babcock & Wilcox, Foster Wheeler North America, and the EERC.

Oxygen firing in circulating fluid-bed boilers may have an advantage over pc firing in that a significant degree of temperature control can be achieved by recirculating solids. Lower flue gas recycle would reduce parasitic power load for fans. In addition, higher O_2 concentrations may be possible, resulting in a smaller boiler island size and reduced capital cost. Development issues center around continuous solids recirculation. Currently, testing is at the large pilot scale, with development efforts being conducted by ALSTOM Power, ABB Lummus Global, Praxair, and Parsons Energy.

2.3 Postcombustion Removal

Removal of CO_2 from low-pressure (<2 psig), low-CO_2-concentration (<15 vol%) flue gases takes place following the pollution control devices, as shown in the schematic in Figure 3.

2.4 Processes Used to Capture CO_2

Several types of processes have been or are being developed to separate and remove CO_2 from a flue gas stream. Figure 4 summarizes the basic types, which are discussed in the subsections that follow.

2.4.1 Absorption

Absorption processes are commonly used in commercial plants to remove CO_2 from mixed-gas streams over a wide range of pressures and CO_2 concentrations. Two types of solvents are typically used for CO_2 removal: physical solvents and chemically reactive solvents. Physical solvents dissolve CO_2 following Henry’s law, but do not react with it. Chemically reactive solvents first dissolve CO_2 and then react with it. Physical solvents are more suitable for mixed-gas streams that are under high pressure. The elevated pressure increases CO_2 solubility, which, in turn, reduces the solvent circulation rate. Pressure does not affect the performance of chemically reactive solvents.

If the mixed-gas stream containing CO_2 is at elevated pressure, the physical solvent can be recovered by flashing off CO_2 at a lower pressure. Chemically reactive solvents require heat to separate the dissolved gas. Commercial experience has shown that the physical solvent process is more economical if the CO_2 partial pressure is above 200 psia. At low inlet CO_2 partial pressure, and where a very low outlet CO_2 concentration is required, chemically reactive solvent processes are more effective.

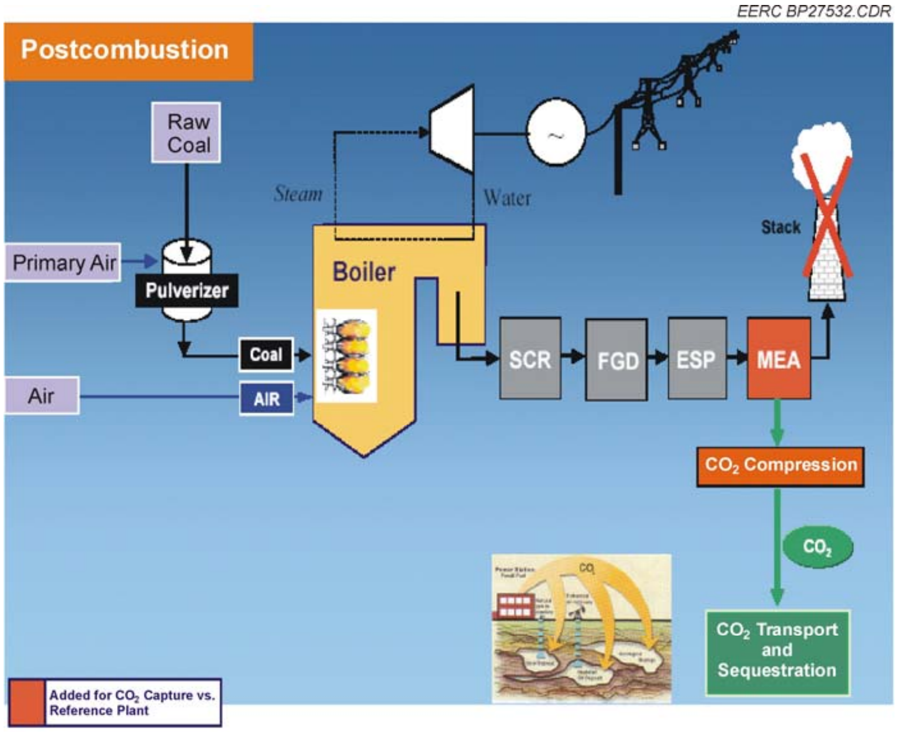

Figure 3. Schematic for postcombustion CO_2 capture (8).

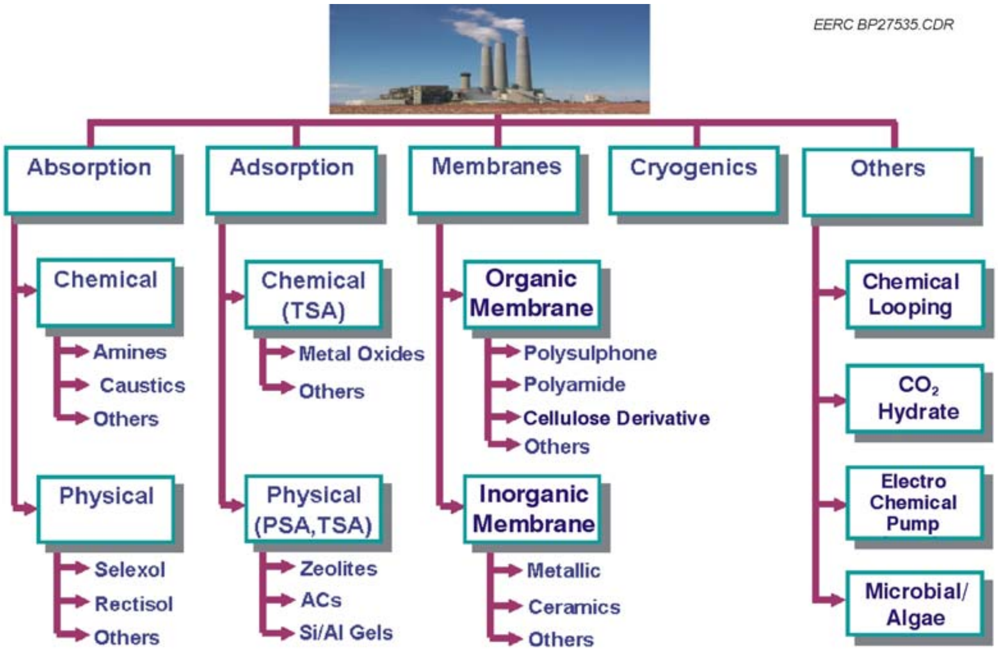

Figure 4. CO_2 capture and separation technology types (8).

Hybrid solvents combine the best characteristics of both chemical and physical solvents and are usually composed of a number of complementary solvents. Work is under way to develop tailor-made complementary solvents where the proportions are varied to suit the application.

2.4.1.1 Chemical Absorption

The most readily available chemical absorption system is amine scrubbing, and it is considered state of the art for fossil fuel-fired systems such as boilers and gas turbines. Alkanolamines are a group of amines used for CO_2 removal that include monoethanolamine (MEA), diethanolamine (DEA), diglycolamine (DGA), diisopropanolamine (DIPA), and triethanolamine (TEA). Of these, MEA is the most alkaline; it has the highest dissociation constant and the highest pH in water solution. The others are progressively less alkaline in the order listed. Other properties that bear on the use of these amines follow in the same order as their alkalinities. The chemical reaction with CO_2 is fastest with MEA and decreases with the others. For these reasons, MEA-based processes are the most popular and are considered the best available control technology (BACT) for the removal of CO_2 from flue gas with low concentration and low partial pressure of CO_2.

Several commercial facilities use MEA-based solvents to capture CO_2 from coal-, fuel oil-, and natural gas-derived flue gas streams for use in the food industry and, in the past, for enhanced oil recovery (EOR). These plants have had capacities in the range of 100 to 1100 tons/day, which is significantly less than the 9000 tons/day produced by a 500-MW coal-fired plant. Commercial providers of MEA technology include Fluor Daniel and ABB Lummus Global. A diagram of a system employing an MEA process for CO_2 capture is presented in Figure 5. In gas turbine combined-cycle systems, as shown in Figure 5, flue gas from the HRSG is cooled to about 110°F with circulating cooling water. Additional cooling is not required in systems employing flue gas desulfurization (FGD). The flue gas is partially compressed to 17.5 psia by a centrifugal blower to overcome the gas path pressure drop. The flue gas enters the absorber base and flows upward countercurrent to the lean MEA solution. CO_2 is removed from the flue gas in the packed-bed absorber column through direct contact with MEA. The CO_2-depleted flue gas is exhausted to the atmosphere. The CO_2-rich solution is heated in a heat exchanger and sent to the stripper unit where low-pressure (LP) steam from the steam turbine crossover provides the thermal energy to liberate the CO_2. The CO_2 vapor is condensed, cooled, and sent to a multistaged compressor where the CO_2 is compressed to a pressure of over 1200 psia. The CO_2-laden stream is dehydrated using glycol or molecular sieve processes. After drying, the CO_2 is ready for transport and sequestration.

The MEA process can achieve recoveries of 85% to 95%, with CO_2 purities over 99 vol%. However, the MEA process also requires a significant amount of power to operate pumps and blowers for gas and solvent circulation. The largest parasitic load to the power cycle is associated with the steam used for solvent regeneration. Energy consumption as steam can be as high as 3.6 to 4.5 million Btu/ton CO_2 recovered. Additional issues with the process are equipment corrosion; solvent degradation caused by the presence of dissolved O_2 and other impurities; or reaction with SO_2, SO_3, and NO_x to produce nonregenerable, heat-stable salts. This requires SO_2

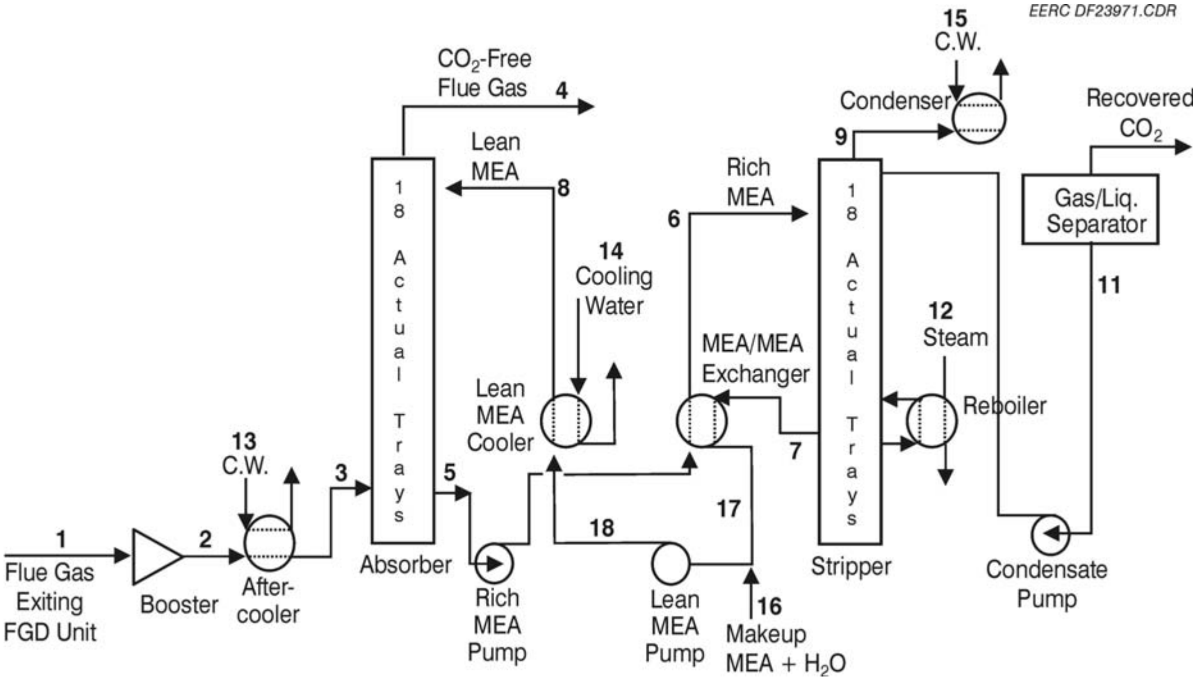

Figure 5. Schematic of an MEA absorption system for capturing CO_2 from flue gas.

levels below 10 ppm, NO_2 levels below 20 ppm, and NO_x below 400 ppm. Solvent degradation and loss also occur during regeneration.

Recent advances in chemical solvents have included the commercial introduction of the KS-family of sterically hindered amines by Mitsubishi Heavy Industries (MHI). Their molecular structure is tailored to enhance reactivity toward a specific gas component, in this instance CO_2. Benefits relative to MEA include higher absorption capacity (only 1 mol of hindered amine is required to react with 1 mol CO_2 compared with 2 mol MEA), 90% less solvent degradation, 20% lower regeneration energy, 15% less power, 40% lower solvent recirculation rates due to higher net absorption capacity, lower regeneration temperature, less corrosion in the presence of dissolved oxygen, and lower chemical additive cost.

Other advanced liquid sorbent systems being developed include:

- A potassium carbonate/piperazine complex (University of Texas at Austin) appears to be an effective additive for improving overall amine solvent performance. The process may permit the use of waste heat because the regeneration temperature is lower (131° vs. 248°F).

- The aqua (aqueous) ammonia process (Powerspan, National Energy Technology Laboratory [NETL]), which has the ability to also capture SO_2, NO_x, HCl, and HF and to produce potentially marketable by-products.

- PSR solvents (University of Regina, Saskatchewan), which are proprietary designer solvents formulated for optimized separation of CO_2 from any gas stream.

- Advanced amine scrubbing (Cansolv Technologies, Inc.), in which a proprietary tertiary amine is utilized.

- Amine-enriched sorbents (NETL).

- Amino acid salt solutions (Norwegian University of Science and Technology) mimic the CO_2 capture by hemoglobin in the blood. The reaction kinetics may be more favorable than traditionally used amines.

These chemical solvent systems are being developed to improve the cost-effectiveness of CO_2 capture through higher CO_2 absorption capacity, faster CO_2 absorption rates (to achieve lower solvent circulation rates and smaller equipment), reduced solvent degradation, less corrosiveness, and lower regeneration energy requirements. Development efforts for these technologies range from bench to pilot scale.

2.4.1.2 Physical Absorption

Physical absorption is primarily used to remove CO_2 from gasification flue gas. These systems include the Rectisol and Selexol processes. The Rectisol process removes CO_2 and H_2S in methanol at –94°F, requiring significant gas cooling and reheating. With respect to potential future requirements for high (>90%) CO_2 recovery during gasification, the double-stage Selexol process, in which desulfurization and CO_2 separation are combined, is favored. The double-stage or double-absorber Selexol unit preferentially removes H_2S in one product stream and then removes CO_2 as a second product stream. The synthesis gas enters the first absorber unit at approximately 705 psia and 103°F. In this absorber, H_2S is removed from the fuel gas stream by "loading" the lean Selexol solvent with CO_2. The CO_2-saturated solvent preferentially removes H_2S. The rich solution is regenerated in a stripper by heating. The stripper acid gas stream, consisting of 34% H_2S and 58% CO_2 and water, is then sent to a Claus sulfur removal unit.

Following processing in the Claus unit, cleaned fuel gas from the first absorber is cooled and routed to the second absorber unit. In this absorber, the fuel gas is contacted with lean solvent. The solvent removes approximately 97% of the CO_2 from the fuel gas stream. The fuel gas from the second absorber is warmed and humidified in the fuel gas saturator, reheated and expanded, and then sent to the burner of the combustion turbine. CO_2 is flashed from the rich solution and is then ready for dehydration and compression to pipeline-ready conditions.

Dry, regenerable, solid sorbents are also being developed for postcombustion CO_2 capture in both low- and elevated-temperature flue gas. With these sorbents, essentially pure CO_2 (>99%) is recovered owing to selective absorption of CO_2. Dry regenerable solid sorbent systems under development include the following:

- Alkali carbonate system (Research Triangle Institute [RTI]).

- Warm-gas sodium-based solid sorbents (NETL) have been developed for PSA/temperature swing adsorption (TSA) application for removal from integrated gasification combined-cycle (IGCC) syngas streams. They rely on chemical reaction for CO_2 capture. Regeneration temperatures are currently too high at 1292°F.

- Ionic liquids (University of Notre Dame, Sachem Inc., Merck) can dissolve gaseous CO_2 and are stable at temperatures up to several hundred degrees Celsius. Little heat is required for CO_2 recovery.

The first process requires a multiple reactor system, with absorption occurring in one reactor and transfer of the loaded sorbent to a second reactor for regeneration and release of CO_2. The second process involves cyclic use of multiple beds, similar to PSA/TSA. Because these are dry systems, there is no need to heat and cool large amounts of water, as required in an MEA system, which leads to lower regeneration energy requirements. Another advantage is the higher contact area for CO_2 absorption. Development efforts for these technologies are currently at the bench scale.

2.4.2 Adsorption

Gas–solid adsorption systems that may be applicable to removal of CO_2 from mixed-gas streams employ adsorbent beds of alumina, zeolite, or activated carbon. Other solid materials used commercially in gas separation processes are alumina gel and silica gel, although processes using these gels are a hybrid of adsorption and absorption.

Four methods are used commercially for regeneration. PSA involves raising and lowering the pressure in the bed to preferentially capture and release the gases to be separated. Technologies requiring a vacuum for regeneration are called vacuum pressure swing adsorption (VPSA) units. PSA and VPSA regeneration cycles are relatively short and are typically measured in seconds. TSA employs high-temperature regeneration gas to drive off trapped gases. TSA regeneration cycles are quite long (measured in hours) and require larger quantities of adsorbent than PSA systems. A third-regeneration method employs a stream of fluid that does not contain any of the trapped gas to “wash” the bed. The fourth method uses a gas stream that contains a material that can displace the trapped gas from the bed and is essentially a chromatographic procedure.

Most commercial units use either PSA-type regeneration or a combined thermal swing/wash method that regenerates at reduced pressure, known as thermal swing. PSA technology is used for drying air, hydrogen purification refineries, *n*-paraffin removal, and small- to medium-scale air fractionation. Depending on the feed gas and the species to be adsorbed, two vessels are filled with an adsorbent such as silica gel, molecular sieves, or molecular sieve carbon. One vessel serves as an adsorbing bed, with the feed entering at elevated pressure. When the bed is saturated, the feed is switched to the second vessel. Pressure in the first (spent) vessel is lowered to release the adsorbed species. The adsorbent in the vessel is regenerated, and the vessel is pressurized to make it ready for another cycle. The process is repeated in the second vessel. Similar to the absorption process, the adsorption can be primarily chemical or physical with physical adsorption being the less energy-intensive to reverse. These processes compete with cryogenic air separation units in applications requiring high-purity products, where the number of stages and the recycle flow rates increase to such an extent that the adsorption processes cannot successfully compete.

Other adsorption techniques include the following:

- Electrical swing adsorption (Oak Ridge National Laboratory) uses carbon-bonded activated carbon fiber as adsorption material. Adsorbed gas is removed by a low-energy electric current.

- Sorption-enhanced water–gas shift process (Air Products), in which a water–gas shift catalyst is combined with CO_2-selective hydrotalcite adsorbent. Multiple adiabatic fixed beds are used for cyclic reaction/adsorption and regeneration.

- Metal organic frameworks (UOP, University of Michigan, Northwestern University) consist of large molecules with engineered macromolecular cavities that can adsorb CO_2. A high storage density is possible, and low heat is required for CO_2 recovery.

2.4.3 Membranes

Although very effective for CO_2 removal, current solvent absorption methods produce system efficiency losses principally because of the need to cool the raw gas to near or below ambient temperature. To improve overall power generation efficiency, new CO_2 separation approaches are being developed for both gasification and reforming applications. Most of these technologies are based on selective membrane systems.

Gas separation membranes use partial pressure as the driving force for separation and, consequently, will be most effective at high CO_2 concentrations and pressure. Differences in physical or chemical interaction between the components present in a gas mixture with the membrane material cause one component to permeate through the membrane faster than the other component. The gas component dissolves into the membrane material and diffuses through it to the other side. The membrane divides the feed gas stream into the permeate stream and the retentate stream. Ideally, the permeate stream would require little recompression for utilization.

The quality of the separation is determined by the membrane selectivity and by two process parameters: 1) the ratio of the permeate flow to the feed flow and 2) the ratio of permeate pressure to the feed pressure. Depending upon the selectivity of the membrane, a high-purity CO_2 product may require a large number of stages, leading to increased recompression and capital costs. Membrane separation often competes with cryogenic separation and PSA when medium quantities of low-purity product gas are required. Membrane separation technology is currently better suited to treatment of mixed-gas streams fed from a high-pressure source, such as natural gas processing.

Gas absorption membranes are used as contacting devices between a gas flow and a liquid flow. The presence of an absorption liquid on one side of the membrane selectively removes certain components from a gas stream on the other side of the membrane. In effect, the absorption liquid increases the driving force across the membrane because the partial pressure of the absorbed gas on the liquid side is essentially zero. In contrast with gas separation membranes, it is not essential that the absorption membrane be selective as its purpose is solely to provide a contacting area without mixing gas and absorption liquid flow. The selectivity of the process is derived from the absorbing liquid.

Removal of flue gas components such as SO_2 or CO_2 is achieved through the use of porous hydrophobic membranes in combination with suitable absorption liquids, such as sulfite, carbonate, or amine solutions. For example, CO_2 is removed from flue gas with the aid of gas absorption membranes used in combination with MEA.

New systems under development include the following:

- Enzymatic liquid membranes (Carbozyme, CO_2 source) are enzyme-based systems that achieve CO_2 capture and release by mimicking the mammalian respiratory mechanism. Its fast kinetics can lower system size and cost.

- CO_2 selective membrane (Media and Process Technology, University of Southern California), a membrane reactor that combines water–gas shift with CO_2 removal. It employs a tubular ceramic membrane, permeable only to CO_2, inside a water–gas shift reactor.

- Membrane water–gas shift reactor (Eltron Research/SOFCo/Chevron Texaco) is a catalytic membrane reactor (CMR) that utilizes oxygen transport membrane technology to facilitate in situ partial oxidation reforming. Syngas passes to a dense metal alloy membrane reactor to facilitate selective permeation of H_2 and enhanced shift. H_2 permeabilities are one order of magnitude higher than palladium and two orders of magnitude less expensive. This process requires sweet syngas, however.

- Hydrogen membrane reformer (Norsk Hydro, SINTEF, and UiO) is a two-reactor process that combines reforming, water–gas shift reaction, and H_2 separation. It utilizes a dense, mixed conducting membrane (MCM). Since the transport process is based on ion diffusion, the selectivity of the membrane is infinite as long as the membrane is gas impervious (barring any defects).

- Palladium membrane reactor (NETL). This reactor system combines a palladium-based membrane with the water–gas shift reaction. The high temperature (1652°F) and pressure of operation and the catalytic effect of the membrane eliminate the need for a separate water–gas shift catalyst. A sulfur-tolerant membrane is possible.

- Hybrid alumina/organosilane membrane (NETL). In this system, organic molecules are grafted onto a substrate surface to attain higher selectivity toward CO_2 permeation.

- Thermally optimized polymer membrane (Los Alamos National Laboratory [LANL], Idaho National Laboratory, Pall Corporation, University of Colorado, Shell Oil Company), in which polymer-based membranes exhibit high selectivity because of size-based exclusion and solubility variances of molecules within the polymer matrix. Polymer membranes have been commercially successful for a number of industrial applications. The intent of additional research is to increase the temperature of application.

- Inorganic nanoporous membrane (Oak Ridge National Laboratory [ORNL]) with pore sizes of less than 1 nm. Composite membranes made of a ~2-µm membrane (e.g., alumina) layer on a ~450-µm porous support structure can allow operation at temperatures up to 1852°F.

- Molecular gate membrane (Research Institute of Innovative Technology for the Earth [RITE]), a cardo-polyimide membrane that is selective to CO_2 permeation.

- Kvaerner hybrid membrane absorption system (Kvaerner Process Systems), a gas–liquid membrane contactor that replaces a traditional absorber. CO_2 diffuses through a microporous, hydrophobic solid membrane into liquid flow. The solvent, rather than the membrane, provides the selectivity. Compared to a conventional absorber, it weighs 70% less and has a 65% smaller footprint.

2.4.4 Cryogenics

CO_2 can be separated from a mixed-gas stream by liquefaction when the CO_2 concentration is sufficient. CO_2 can be liquefied at any temperature between its triple point 70°F and its critical point 88°F by compressing it to the corresponding liquefaction pressure and removing the heat of compression and condensation. There are three common commercial liquefaction processes.

In the first process, the CO_2 is liquefied near the critical temperature, and water is used for cooling. This process requires compression of the CO_2 gas to about 1100 psia. A second liquefaction process operates at temperatures from 10° to 70°F and with a liquefaction pressure of about 250 to 350 psia. This process requires dehydration of the feed stream with an activated alumina or silica gel dryer and distillation of the condensate in a stripping column. The third cryogenic process cools the mixed-gas stream to a temperature sufficiently low to condense CO_2 out of the gas phase. This method is also used to remove vapors of organic compounds from vent gases and for other operations.

2.4.5 Technologies Related to Oxygen Combustion

As previously mentioned, the high cost of oxygen separation is a major issue with oxygen combustion. State-of-the-art cryogenic distillation air separation has little room for improvement or cost reduction. Current development activities are centered on ion transport membranes. These are complex crystalline structures with oxygen ion vacancies onto which oxygen adsorbs and decomposes into ions. The ions are transported through the membrane by sequential occupation of oxygen ion vacancies with the ion transport balanced by the counterflow of electrons. Oxygen partial pressure provides the driving force, which requires high-pressure air at temperatures above 1292°F. Barring the presence of defects, the membrane is selective to oxygen transport only.

The ion transport membranes can theoretically integrate high-temperature oxygen separation from air with the combustion process, leading to a significant reduction in parasitic power as well as lower cost for O_2 production. Development issues include materials of construction, integration with or into the boiler, control of wall temperature (as a consequence of combustion reaction), and carbon formation. Developers and systems include Praxair and ALSTOM Power (oxygen transport membrane [OTM]), and Air Products (ion transport membrane [ITM]).

Concepts being developed that utilize ion transport membranes for oxygen separation include the following:

- Advanced zero emission power (AZEP) process (ALSTOM Power, Norsk Hydro), which is utilized with conventional gas turbines. Air from the compressor is supplied to a new mixed conducting membrane (MCM) reactor. The reactor combines O_2 separation, combustion, and heat transfer. Preliminary evaluations show a 2% loss in plant efficiency for separation vs. a 10% loss with flue gas CO_2 separation.

- Integration into a fired boiler (Praxair) in which an OTM is incorporated directly into the boiler. It can be utilized with gaseous or liquid fuel.

- Utilization with a circulating fluid-bed (CFB) or circulating moving-bed (CMB) boiler (ALSTOM Power). In this case, the OTM stands alone but is thermally integrated with the boiler. It requires a high-temperature air source and is heated by in-bed heat exchange of CFB or CMB.

2.4.6 Other Processes

Other advanced processes and/or unconventional systems are being developed with the intent of improving efficiencies and lowering the cost to capture and purify CO_2. These include the following:

- Regenerative carbonate process (ALSTOM Power), which utilizes a recirculating stream of lime (CaO) to capture CO_2 during combustion. Calcium carbonate is regenerated in a calciner to liberate pure CO_2. In this process, there is no thermodynamic efficiency loss for CO_2 capture.

- Chemical looping gasification (ALSTOM Power). The intent of this process is to produce a nearly pure CO_2 stream and a medium-Btu gas (>90% H_2) after CO_2 separation. The high energy and cost penalty associated with O_2 separation is avoided (it is similar in concept to chemical-looping combustion). It uses two separate chemical loops: one for oxygen transfer and one for CO_2 capture.

- ZEC technology (ZECA Corporation), hydrogasification of carbon-based fuel to H_2 with CO_2 capture in a carbonate cycle. It would achieve permanent sequestration of CO_2 through mineral carbonization. The ultimate embodiment of the process is the high-efficiency (70% to 75%) conversion of H_2 fuel gas to electricity through application of a coal-compatible fuel cell (CCFC).

- Unmixed fuel processor (UFP) (GE Global Research [GEGR]) is a gasification process developed to convert coal, steam, and air into hydrogen, sequestration-ready CO_2, and a low-quality, high-temperature air stream used for power production in a gas turbine. Regenerable oxygen transfer material is used to provide oxygen for the process; no external ASU is required.

- CO_2 hydrate (SIMTECHE, Nexant, LANL) is a below-ambient-temperature, high-pressure, aqueous-based process that captures CO_2 from syngas through the formation of CO_2 hydrates. H_2 acts as an inert and is not retained in the hydrate crystal. The CO_2 is recovered from the hydrate slurry by heating and reducing pressure.

- CO_2 hybrid process (Foster Wheeler North America). This process combines oxygen-blown partial gasification with oxygen combustion of syngas in a gas turbine. The gas turbine exhaust provides sensible heat and oxygen for char combustion to produce steam for partial gasification. Flue gas from the char combustion contains all of the CO_2 from the process; recovery is accomplished by compression and flash of noncondensables.

- Chemical looping combustion or sorbent energy transfer system (TDA Research, ALSTOM Power, Chalmers University). In this technology, separation of CO_2 occurs during combustion, and no energy is expended for CO_2 separation. There is no direct contact of fuel with air, and no ASU is required. An oxygen carrier transfers oxygen from the combustion air to the fuel. The net chemical reaction and heat release are equivalent to that of conventional combustion. The process is currently applicable only to gaseous or liquid fuels unless the solid fuel is first gasified in O_2.

- Water cycle (Clean Energy Systems) is based on a high-temperature/pressure aerospace-derived gas generator operating at 1500 psi and 3000°F. The fuel is fired stoichiometrically with oxygen, and water is injected to control temperature and protect gas generator components. The working medium is a high-pressure, high-temperature steam–CO_2 mixture comprising 90% steam and 10% CO_2.

- Graz cycle (Institute for Thermal Turbomachinery and Machine Dynamics). As with the water cycle, it uses a 25:75 steam–CO_2 mixture as the working fluid. It combines the gas turbine cycle with the steam cycle to improve efficiency. Gaseous fuel is reacted with stoichiometric oxygen in the combustor at 580 psi with steam (as opposed to water) injected for temperature control.

- MATIANT cycle (Institute of Mechanical Engineering, University of Liege [Belgium]) combines a Brayton-like cycle (in which CO_2 serves as the principal working fluid) with a steam cycle.

3.0 EVALUATION OF CO_2 CAPTURE TECHNOLOGIES FOR NORTH DAKOTA POWER PLANTS

Power plants in North Dakota will typically have CO_2 concentrations in the flue gas ranging from 10% to 15%, which is typical for pc-fired power plants. CO_2 capture becomes challenging at these relatively low CO_2 concentrations. Two technologies were chosen for evaluation for a typical North Dakota power plant: chemical absorption (amine scrubbing) and oxyfuel combustion. A model was created in Excel that can be tailored to different North Dakota power plants for a given set of inputs. The major inputs necessary for the model are plant size, capacity factor, net heat rate, and coal properties such as heating value and ultimate analysis values on an as-received basis. Default values can be used when inputs are not known. The following section will provide a detailed description of the two chosen technologies.

3.1 Amine Scrubbing

Amine-based CO_2 absorption has been studied in the past and identified as one of the most suitable means for removing CO_2 from combustion-based power plants for the following reasons:

- The systems are effective for dilute CO_2 streams, such as are typically found in North Dakota lignite-fired facilities.
- The technology is proven and commercially available.
- The units are operated at standard temperature and pressures similar to other pollution control devices currently employed at power plants.
- A current worldwide effort is being undertaken to improve amine systems because of their potential role for wide-scale CO_2 capture; therefore, future benefits from technology advances are anticipated.

Amine-based absorption/stripping has been around for several decades as a commercial technology for CO_2 removal from natural gas and hydrogen. The amine process was first patented by R.R. Bottom in 1930 for acidic gas treatment. Throughout the years, the amine-based gas treatment process has remained relatively unchanged. The concept of removing or capturing CO_2 from flue gas streams started back in the 1970s as a possible economical source of CO_2, mainly for EOR operations. Today, about 80% of CO_2 production is used for EOR applications, most of which is obtained from natural CO_2 domes (9). CO_2 is also produced for several other industrial applications, including carbonation of brine, dry ice production, urea production, and in beverages.

Several commercial CO_2 plants were constructed in the late 1970s and early 1980s in the United States (10, 11). Although some of these plants are still in operation today, all of them are much smaller than a typical power plant in terms of tonnage of CO_2 handled or produced. Once the CO_2 is captured, it has to be securely stored (sequestered) to prevent it from entering the atmosphere, unless an application is identified for the captured CO_2. Very little experience exists with the sequestration of CO_2 other than in EOR applications. The first commercial CO_2 sequestration technology started in 1996 in response to a Norwegian carbon tax. The facility, Statoil, has been storing approximately one million tonnes of CO_2 per year from the Sleipner West gas field into a sandstone aquifer 1000 meters beneath the North Sea (12, 13). Statoil's operations have been monitored closely by the research community in order to enhance the understanding of geological sequestration.

3.1.1 Commercial Processes

Two main MEA-based systems are available for commercial CO_2 capture plants: the Fluor Daniel Econamine FG process and the ABB Lummus Crest MEA process (14).

3.1.1.1 Fluor Daniel Econamine FG Process

The process was developed as an inhibited 30 wt% MEA process by Dow Chemical and Union Carbide. The Econamine FG process uses a 30% w/w MEA solution with an oxygen inhibitor. The oxygen inhibitor helps in two ways: reduced sorbent degradation and equipment corrosion. The process was known as the GAS/SEC FT-1 process from the 1970s until 1989 when Dow Chemical sold the process to Fluor Daniel, Inc. In the 1970s and 1980s, the process was mainly used to capture CO_2 from flue gases for the EOR market. When Fluor Daniel took over the process in 1989, it renamed the technology Econamine FG. Sixteen commercial plants have been built, including nine large plants of 60 tons/day or more (seven of which are still operational) and seven skid-mounted plants (9). Fluor Daniel licensed three of the nine commercial plants that range in size from 90 to 320 tonnes/day. Prior to this, Dow licensed six commercial plants ranging from 6–1000 tonnes/day. In comparison, a 500-MW gross coal-fired power plant would need to capture 9000 tonnes/day.

At this time, all of the large commercial units treat flue gas from the combustion of natural gas, with the exception of the Sumitomo Plant, which fires a variety of fuels. The Econamine FG process has also been demonstrated at three pilot-scale plants that use a coal-derived flue gas. These plants process 2, 4, and 4.5-tonnes CO_2/day (15). Recently, Fluor Daniel announced an improved version of the Econamine FG process called "Fluor Econamine FG Plus," which addresses some of the issues in the Econamine FG process (16). No commercial plants yet exist that employ the Fluor Econamine FG Plus.

3.1.1.2 Kerr-McGee/ABB Lummus Amine Process

Kerr-McGee started its 800-tonne/day CO_2 capture unit, the Trona unit, in 1978. The Trona unit has been fed flue gases from boilers firing natural gas, coal, and coke. Reliability and cost-effectiveness of the installation were improved during the first several years of operation. In 1990, Kerr-McGee partnered with Lummus Global, whereby Lummus gained worldwide exclusive marketing rights to Kerr-McGee's CO_2 capture technology. Three units have been licensed using this technology to Applied Energy System, Poteau, Oklahoma, and Soda Ash Botswana, Pty. Ltd., Sue Pan, Botswana. Applied Energy System is a 300-MW coal-fired cogeneration plant that has been capturing and producing a 200-tpd food-grade liquid CO_2 product since 1991. Applied Energy Systems installed a second unit in 1997 with a capacity of 150 tpd. Soda Ash Botswana is a 300-tpd CO_2 unit which has been in operation since 1991.

The Kerr-McGee/ABB Lummus process uses a 15%–20% w/w MEA solution without any inhibitor (17). The technology is effective at capturing greater than 96% of the CO_2 from flue gases, but the lower sorbent concentration leads to economic disadvantages. The lower sorbent concentration requires larger equipment because of its higher flow rates and higher energy requirements because of the increased amounts of dilution water per unit of sorbent. One advantage of this system is that it has already been proven at commercial scale for coal-fired flue gas applications.

3.1.2 Process Chemistry

CO_2, MEA, and water (H_2O) are the three main compounds that are active in an amine-scrubbing system. The following equilibrium reactions occur in the bulk of the liquid (18).

Water hydrolysis: $2H_2O \leftrightarrow H_3O^+ + OH^-$
Bicarbonate formation: $2H_2O + CO_2 \leftrightarrow H_3O^+ + HCO_3^-$
Carbonate formation: $H_2O + HCO_3^- \leftrightarrow H_3O^+ + CO_3^{2-}$
Amine protonation: $H_3O^+ + R - NH_2 \leftrightarrow H_2O + R - NH_3^+$
Carbamate formation: $R - NH_3 + HCO_3^- \leftrightarrow H_2O + R - NH - COO^-$

where MEA is represented by R-NH_2 and “R” stands for HO–CH_2–CH_2.

The process chemistry is complex, but the main reactions taking place are (19):

CO_2 absorption: $2R - NH_2 + CO_2 \rightarrow R - NH_3^+ + R - NH - COO^-$
MEA regeneration: $R - NH - COO^- + R - NH_3^+ + (heat) \rightarrow CO_2 + 2R - NH_2$

Pure MEA is an unhindered amine that forms a weakly bonded intermediate ion called carbamate, which is fairly stable. For every mole of amine, one-half mole of CO_2 is absorbed, as shown in the above CO_2 absorption equation. Upon the application of heat, the carbamate dissociates to give back CO_2 and amine sorbent, as shown in the MEA regeneration equation above. Since the carbamate is fairly stable, it takes a substantial amount of energy to break the bonds and regenerate the sorbent. The theoretical minimum heat requirement to regenerate the MEA is about 1900 kJ/kg CO_2. The actual heat requirement is greater than double this theoretical minimum.

Despite the use of inhibitors and dilution with water, a small quantity of MEA is lost through various unwanted reactions. Two main side reactions occur: the polymerization reaction that forms long-chained compounds and the oxidation reaction forming organic acids and liberating ammonia. Appropriate measures must be taken to avoid accumulation of the unwanted chemical species in the circulating sorbent. Flue gas impurities (acid gases) are another potential source of sorbent loss, especially for coal-fired flue gases. Therefore, very low concentrations of these gases, on the order of 10 ppm, are desirable to avoid excessive loss of sorbent. The problem is especially acute for SO_2 because its concentration in flue gas is typically

700 to 2500 ppm at coal-fired plants. NO_x is less of a problem because only NO_2 (which makes up only about 5% of the total NO_x) reacts with MEA.

3.1.3 Process Description

The technology chosen for the analysis in this project was Fluor Daniel's Econamine FG process. This system was chosen because of the availability of data and literature sources that show that it is currently the most proven of the two technologies. The main tool used for analysis for this project was the Integrated Environmental Control Model (IECM) developed by Carnegie Mellon University with support from DOE NETL. IECM is a tool for calculating the performance, emissions, and cost of a fossil-fueled power plant. More details of IECM can be found in Section 4.0. Table 2 shows the major equipment required for the amine-based CO_2 capture system.

The amine-scrubbing unit would be installed downstream of any existing pollution control devices, such as those used for particulate, NO_x, and SO_x removal. It must be noted that in order for this system to operate with minimum solvent degradation, SO_2 control is a must. Currently, in North Dakota, about a third of the megawatts are generated at plants that do not currently control their SO_2 emissions. In these cases, an SO_2 control device (such as a spray dryer or wet FGD) must be installed for amine scrubbing to be feasible for capturing CO_2. Although an SO_2 control device is necessary for amine scrubbing, upcoming regulations will probably require the installation of these devices prior to installing a CO_2 capture technology to meet SO_2 emission requirements. The maximum allowable amount of SO_2 that can be present in the flue gas is 10 ppmv, which may require several plants to upgrade their existing SO_2 control devices. Figure 6 shows the breakdown of SO_2 pollution control devices for North Dakota power plants based on megawatts of production. A basic block flow diagram demonstrating the installation of an amine-scrubbing system to an existing power plant can be seen in Figure 7.

The CO_2 capture plant, including the amine-scrubbing unit, consists of four process modules: flue gas pretreatment, absorption, stripping, and CO_2 compression and drying. In the flue gas pretreatment section, the flue gas is cooled and conditioned before it enters the absorber; CO_2 is removed in the absorber by contacting the flue gas countercurrently with an MEA solution. Once the CO_2 is absorbed in the MEA, the CO_2-rich solvent is sent to the stripping section where the CO_2 is removed from the solvent by the addition of heat. The regenerated solvent is sent back to the absorber, while the purified CO_2 stream is sent to the compression and liquefaction unit. Here, the CO_2 is compressed and then dried in the final step, at which point it is ready to transport.

The amine-based CO_2 capture facility may need to consist of several trains to handle the large flow rates from the power plant. A more detailed description of the process follows. Figure 5 contains a process flow diagram for the amine-scrubbing system. A study done by ALSTOM Power Inc. determined that an MEA-based CO_2 capture facility for a ~500-MW power plant required a total plot plan area of 7 acres (20).

Table 2. Equipment Necessary for an Amine-Based CO_2 Capture System

Direct-Contact Flue Gas Cooler	Propane Refrig. Subcooler	Wash Water Pump
CO_2 Absorber	CO_2 Compressor First-Stage Air Cooler	Direct-Contact Cooler Water Pump
Solvent Stripper	CO_2 Compressor Second-Stage Air Cooler	Rich-Solvent Pump
Solvent Stripper Reboiler	CO_2 Compressor Third-Stage Air Cooler	Lean-Solvent Pump
Solvent Stripper Reclaimer	Solvent Stripper Bottoms Cooler	Solvent Stripper Reflux Pump
Solvent Reclaimer Effluent Cooler	Solvent Stripper Reflux Drum	Filter Circ. Pump
Solvent Stripper Condenser	CO_2 Compressor Second-Stage Suction Drum	LP Condensate Booster Pump
CO_2 Compr. First-Stage Aftercooler	CO_2 Compressor Third-Stage Suction Drum	CO_2 Pipeline Pump
CO_2 Compr. Second-Stage Aftercooler	Liquid CO_2 Surge Drum	Caustic Metering Pump
CO_2 Compr. Third-Stage Aftercooler	CO_2 Compressor Third-Stage Discharge Knockout Drum	CO_2 Compressor (motor driven)
CO_2 Condenser	Propane Refrig. Surge Drum	Propane Refrig. Compressor
Direct-Contact Flue Gas Water Clr.	Propane Refrig. Suction Scrubber	LP Steam Turbine/Generator
Rich/Lean Solvent Exchanger	Caustic Day Tank	Corrosion Inhibitor Package
Lean Solvent Cooler	DCC Water Filter	Solvent Filter Package
Propane Refrig. Condenser		CO_2 Dryer Package

3.1.3.1 Flue Gas Pretreatment

Flue gas from the desulfurization unit flows through a motor-driven fan in order to increase the pressure to 1.5 psig, enough to overcome the pressure drop through the direct cooler and absorber. The flue gases coming from the power plant can be very hot and may range from as low as 60°C (140°F) in the case of coal-fired plants with wet scrubbers to more than 550°C (1022°F) in the case of a natural gas-fired, simple-cycle power plant. Typical coal-fired power plants without scrubbers have flue gas temperatures of 150°–200°C (300°–400°F). The amine ystem requires flue gas temperatures of about 45°–50°C (104°–122°F) in order to improve the absorption of CO_2 into the amine sorbent, to minimize the sorbent loss, and to avoid an excessive

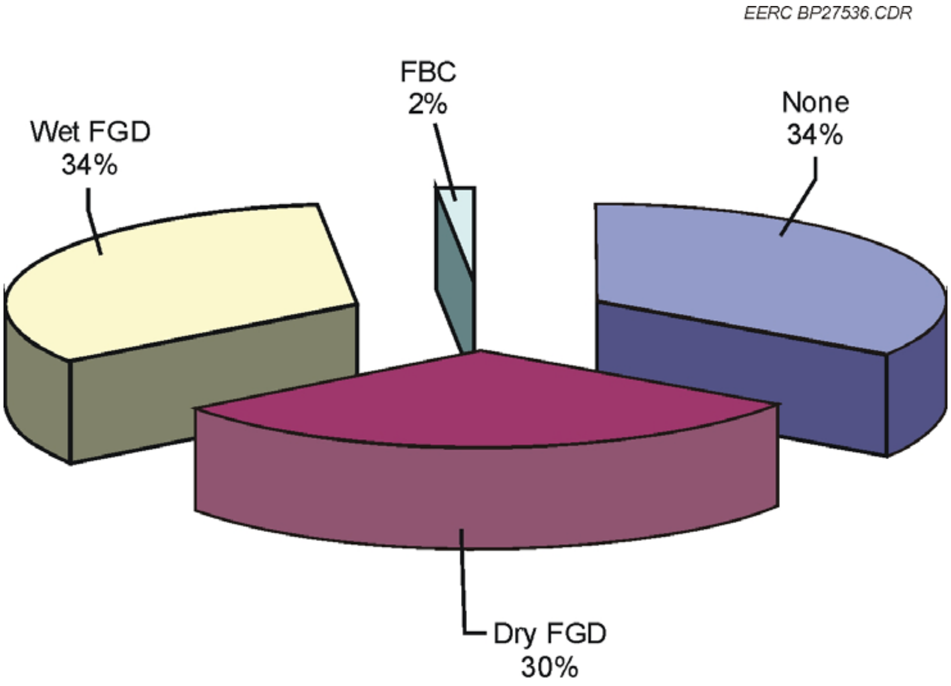

Figure 6. The breakdown of SO_x control technologies for North Dakota power plants based on megawatts of production.

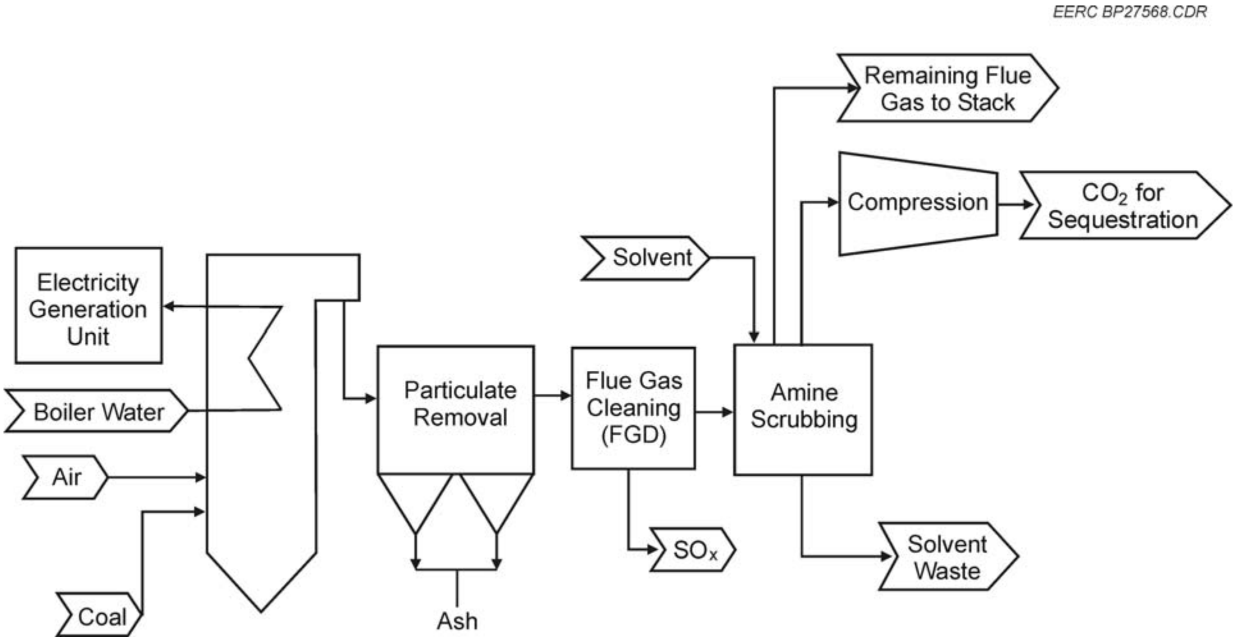

Figure 7. Simple block flow diagram of a coal-fired utility with an amine-based CO_2 capture system.

loss of moisture with the exhaust gases. The absorption process is exothermic and, is, therefore favored by low temperatures. In cases where coal-fired plants are equipped with a wet scrubber, an additional cooler may not be necessary as the scrubber helps in reducing the temperatures.

3.1.3.2 Absorber

Cooled flue gas enters the bottom of the CO_2 absorber and flows upward countercurrently to a stream of 30 wt% MEA solution. The lean MEA enters the top of the column and heats up gradually as it absorbs more and more CO_2 and gains about 20°–30°F with 90%–95% capture. Typical CO_2 loading for lean MEA is 0.2–0.22 mol CO_2/mol MEA. The CO_2-rich MEA leaving the bottom of the column has a CO_2 loading of approximately 0.44 mol CO_2/mol MEA. The CO_2 absorber can be a plate-type column or a packed tower that contains two beds of structured packing and a third bed, usually called the wash zone, at the top of the column. Most of the CO_2 absorbers are packed columns using some kind of polymer-based packing to provide a large interfacial area.

The CO_2-rich solvent exits the bottom of the absorber column and flows through a rich/lean cross heat exchanger. The rich solvent must be heated in order to strip off the CO_2 and regenerate the solvent. The regenerated, or lean, solvent coming from the stripper must be cooled down before it can be circulated back to the absorber column. Therefore, these two streams are passed through a cross heat exchanger where the rich sorbent is heated and the lean sorbent is cooled. This helps to recover some of the energy used to strip the CO_2 from the solvent, thus minimizing the stripper energy requirements.

In this process module, a sorbent-processing area is necessary. The regenerated sorbent needs to be further cooled to an acceptable level of about 40°C after passing through the rich/lean heat exchanger. To make up for solvent losses, a small quantity of fresh MEA must be added to the sorbent stream. The sorbent-processing area, therefore, essentially consists of a sorbent cooler, an MEA storage tank, and a mixer.

3.1.3.3 Stripping (regeneration)

The stripping process module contains all of the equipment necessary for regenerating the sorbent and stripping the CO_2 and consists of a stripping column, reboiler and condenser, reflux drum, steam extractor, and MEA reclaimer. This portion of the process begins as the rich solvent enters near the top of the column. Once in the column, the weak intermediate compound that is formed between the MEA-based sorbent and the dissolved CO_2 (i.e., carbamate) is broken down by the addition of heat, separating the CO_2 from the sorbent. As the solvent flows downward, the hot vapors from the bottom reboiler strip the CO_2 from the solution. Stripping is completed in the reboiler with the addition of more heat. The main drawback of using MEA is that the stability of the carbamate ion requires more heat for the regeneration of the sorbent.

The hot vapors that exit the top of the stripper contain CO_2, water, and solvent. The overhead vapors are cooled in a cold-water condenser where most of the water and solvent vapors condense, but the CO_2 does not. The condensed liquid and gaseous CO_2 are separated in a reflux drum. The CO_2 stream continues on to the CO_2 purification system, while the liquid is returned to the top bed of the stripper.

In coal-fired power plant retrofit cases, a part of the low-pressure (LP)/intermediate-pressure (IP) steam has to be diverted for use in the reboiler for sorbent regeneration. This steam is obtained from the steam turbines by steam extractors.

Acid gases such as SO_2, SO_3, NO_2, and HCl in the flue gas form compounds with the MEA solvent solution that cannot be removed by the addition of heat in the reboiler. These materials are referred to as heat-stable salts (HSS). In order to avoid accumulation of HSS, a small slipstream of the lean solvent from the bottom of the stripper is fed to the MEA reclaimer. The MEA reclaimer is a heat exchanger which vaporizes the free MEA, leaving the high boiling nonvolatile impurities. The reclaimer restores the MEA’s usefulness by removing the impurities such as HSS, suspended solids, acids, and iron products from the solvent solution. Caustic is also added to the MEA reclaimer, freeing the MEA from its bonds with sulfur oxides because of its stronger basic attraction, minimizing MEA loss by allowing more MEA to be vaporized back into the circulating mixture. The reclaimer waste is sent for proper disposal.

3.1.3.4 CO_2 Compression and Drying Unit

The high-purity CO_2 stream from the stripper needs to be prepared for its final use. In order to easily handle the captured CO_2, it must be compressed into liquid form. This is done by using a multistage compressor with interstage cooling. Most of the water is knocked out during compression and is removed with intermediate suction drums. A CO_2 dryer is located after the last stage of compression to meet the water specifications for the CO_2 product. CO_2 is liquefied at about 194 psig and is further pumped to the required pressure. The compression unit yields a final CO_2 product at the specified pressure (typically 2200 psig) that contains acceptable levels of moisture and impurities.

3.1.4 MEA Process Concerns

Although the amine-based absorption process is the most suitable technology currently available for postcombustion CO_2 capture from coal-fired power plant flue gases, several concerns exist: loss of sorbent, energy penalty, corrosion, space constraints, and environmental emissions. These problems are discussed in more detail below.

3.1.4.1 Loss of Sorbent

Sorbent loss can occur throughout the process for a variety of reasons, including degradation, entrainment, vaporization, and mechanical losses (21, 22). Because of flue gas impurities, all of the sorbent that enters the stripper (regenerator) is not regenerated. Sorbent losses due to impurities have already been discussed in detail in previous sections. The MEA reclaimer is the current method used to minimize sorbent losses. Technologies such as electrodialysis are also being proposed for this purpose (23).

3.1.4.2 Energy Penalty

Separation processes in general are very energy-intensive, and amine scrubbing is no exception. Significant energy is required to regenerate the sorbent because of the stability of the carbamate ion and the large quantity of water from dilution of the MEA. Substantial energy is also needed to compress the captured CO_2 to its final product and to meet the transportation requirements. If steam and electricity are extracted internally from a power plant, which is expected for retrofit cases, the large heat and electricity requirement will reduce the net efficiency by derating the plant. In cases of new power plant construction, a bigger power plant will need to be built to produce the energy required by the CO_2 capture facility. If an existing power plant needed or wanted to maintain its power output, an auxiliary boiler and steam turbine would need to be added, thus raising the capital investment necessary for CO_2 capture.

3.1.4.3 Space Constraints

An amine scrubbing facility is estimated to take up a footprint area of ~7 acres based on a 500-MW plant. In North Dakota, this requirement is not a big issue because of the location of the power plants. In other cases, existing power plants may already be low on useable space and would have to buy land at a premium, which would increase costs. In some cases, it may not be possible to acquire the additional land required.

3.1.4.4 Corrosion

Corrosion in an amine-based CO_2 capture system can be a major problem. MEA is a reactive compound. When in solution with water in the presence of oxygen and CO_2, it creates a highly corrosive system. Irreversible side reactions with CO_2 and other flue gas components lead to the formation of various degradation by-products that are associated with increased corrosion in the system. Therefore, controlling corrosion is very important in an amine system where oxygen is present. Several things can be done to limit the rate of corrosion in the system, including reducing the concentrations of MEA and using appropriate materials of construction, corrosion inhibitors, and milder operating conditions (i.e., low temperatures and pressures) (24).

3.1.4.4 Environmental Impacts

Environmental issues may arise from the use of MEA-based CO_2 capture systems, primarily from the spent sorbent slurry, or MEA reclaimer waste, and the emissions of MEA and ammonia carried by the treated flue gas. The MEA reclaimer waste is considered to be a hazardous waste (24). This was further proved by a study performed by NETL, Pittsburgh, which identified chemical species (MEA, ammonia, 3-hydroxyethylamino-N-hydroxyethyl propanamide, 4-hydroxyethyl-2-piperizinone, 2-hydroxyethylamino-N-hydroxyethyl acetamide, and N-acetyletylethanolamine) in the reclaimer waste that are considered to be hazardous (25, 26). Entrainment of MEA with the treated flue gas will be at most a few parts per million and is minimized with the addition of a wash section in the top portion of the absorber column. Although the emission rate of ammonia is higher than for a power plant without an amine-scrubbing system, the ammonia emissions are still much smaller than those from other sources, such as animals and farms. The significance of these environmental impacts is not clear at this time and will need to be considered before the technology can be widely applied (26).

3.2 Oxyfuel Combustion

The second technology chosen to be evaluated is oxyfuel combustion. This technology was not chosen for its commercial availability, but rather because of the ongoing research and its application to CO_2 capture. Oxyfuel combustion for CO_2 capture has been researched extensively in the last few years, and many studies have shown that it is a promising technology option. The CO_2 capture portion of the process is quite simple, possibly involving only condensation of the water and removal of other small quantities of impurities such as particulates and SO_x. Although there are a number of variants in the oxyfuel process, it simply involves modification to familiar pc technology to include oxygen separation; flue gas recycling; and CO_2 compression, transport, and storage.

Throughout time, oxyfuel combustion concepts have been applied to many different processes. Oxyfuel combustion was first introduced in the 1940s as a method of achieving high temperatures for applications such as welding, metal cutting, and flame polishing. In the 1960s, oxyfuel was used to enhance productivity through O_2-enriched combustion in the glass, aluminum, and cement industries. In the 1980s, the concept to use oxyfuel combustion with recycled flue gas for EOR was invented. During the 1980s oxyfuel was also applied to the waste incineration and steel and copper industries as a fuel-saving application. Oxyfuel was employed as a method to reduce NO_x emissions in the 1990s in glass-melting furnaces and small coal-fired boilers. Finally, research since 2000 is centered on using oxyfuel combustion as a method to reduce CO_2 emissions, with likely application to the power generation industry.

3.2.1 Oxyfuel Technology Status

The full-scale application of oxyfuel technology is still under development. However, laboratory and theoretical work have provided an initial understanding of design parameters and operational considerations. There have been a number of investigations at pilot-scale facilities in the United States, Europe, Japan, and Canada. Studies have also assessed the feasibility and economics of both retrofits and new power plants. Some of the conclusions that can be drawn from the findings to date are the following:

- Pilot-scale studies have demonstrated that there are no significant technical barriers to O_2/refinery fuel gas (RFG) firing of pc boilers.

- Typically, the optimum O_2 concentration from the ASU for oxyfuel applications is around 97%–98%, and the optimum recirculation rate is generally around 70%–80%, which yields about 25%–30% O_2 (vol%, wet) in the windbox of the boiler and about 3%–3.5% O_2 (vol%, wet) at the furnace exit/air heater (AH) inlet. At these conditions, flame condition and heat-transfer characteristics reasonably approximate those for air-fired PC boilers.

- O_2/RFG combustion yields significant reductions in NO_x levels, typically 25%–50% lower than for the air-fired case.

- Preliminary cost evaluations indicate that oxyfuel CO_2 capture costs (\$/t CO_2 avoided) and electricity costs (\$/MWh) are comparable with other technologies and are actually lower than the costs of conventional PC with amine-based, postcombustion capture of CO_2.

- Technical challenges include investigation of flame stability, heat transfer, flue gas cleanup requirements, acceptable level of nitrogen and other contaminants for CO_2 compression, and corrosion caused by elevated concentrations of SO_2/SO_3 and H_2O in the flue gas (27).

3.2.2 Process Description

The objective of oxyfuel combustion for CO_2 removal is to produce a flue gas that contains low concentrations of N_2, O_2, Ar and high concentrations of CO_2 and water, simplifying CO_2 capture. The process requires two additional processes to be added to the existing coal-fired power plant: an ASU and a CO_2 capture plant. Figure 8 shows the overall process flow diagram of an oxyfuel combustion facility. CO_2 capture can be achieved by a simpler process than for typical postcombustion capture systems since the flue gas consists predominantly of CO_2. The advantage created by oxyfuel combustion for easier CO_2 capture is somewhat negated by the need for high-purity oxygen during combustion.

The process is quite simple and requires only a few changes to a typical coal-fired utility. Raw coal from the coal bunkers is supplied to the existing coal mills where it is pulverized and transported with recycled flue gas to the boiler. The ASU supplies the oxidant necessary to the existing boiler. Other than through air infiltration, air is not supplied to the boiler. The oxygen produced by the ASU flows through a heat exchanger, where it is preheated by the flue gas leaving the existing air heater. In an oxyfuel system, the air heater is used to preheat recycled flue gas rather than air. The pulverized coal is combusted in the boiler with a mixture of preheated oxygen and preheated recycled flue gas.

The flue gas leaves the boiler and is cooled in a series of heat exchangers. The flue gas leaving the air heater flows through the existing pollution control devices and any other existing heat exchangers and fans necessary for the process. This flue gas stream is split, with approximately one-third of it flowing to the CO_2 separation and compression system. The remaining 70%–80% of the flue gas is recycled back to the boiler. The primary reason for the large recycle stream is to maintain the thermal balance in the boiler and to generate the required boiler performance. The new equipment for this process has been estimated to require 5.5 acres of plot area for a 500-MW facility.

3.2.3 Air Separation Unit

The ASU is the first process needed for an oxyfuel combustion facility. Current processes for oxygen production require between 250 and 450 kWh/ton of O_2 produced. This value is much greater than the theoretical minimum (53 kWh/ton O_2) energy required to separate O_2 from air. This is mostly because of trade-offs in efficiency versus equipment cost (27). Several studies have been performed to determine the most feasible air separation technology for the production of high volumes of high-purity O_2. Results show that cryogenic air separation is the only technology that is currently available to produce the quantity and purity of oxygen needed in the oxyfuel combustion process. Cryogenic air separation is a widely used process and is available in capacities greater than 3500 tpd. Most of the large plants currently in operation today are used in the steel and gasification industries. Even at these large capacities, multiple trains will need to be used in order to produce the oxygen necessary for coal combustion.

A cryogenic ASU is typically a multicolumn, low-temperature distillation (–183°C) unit to produce O_2 from compressed air. This technology can produce high-purity N_2 and Ar as a by-product stream at a relatively low incremental cost, if the market is available (27). Argon recovery is less likely to be economical because of the large scale of O_2 production. If the O_2 purity is greater than 97%, the production of Ar becomes more economical. The process typically produces a gaseous stream of oxygen that is slightly above atmospheric pressure and near ambient temperature. It is possible to produce liquid oxygen from the distillation section upstream of the cryogenic heat exchangers. Using liquid oxygen has advantages to allow fast ramp-up and to protect against short-term outages of the ASU. Unfortunately, producing liquid oxygen from the cryogenic system requires twice the power needed to produce gaseous O_2. Figure 9 is a simplified process flow diagram for a typical cryogenic ASU.

Currently, the largest cryogenic ASU in operation is a 3500-ton O_2/day facility at Rozenburg, The Netherlands. The largest plant under construction to date is a 4300 ton O_2/day (3900 metric tonnes). This plant is called the Long Lake Project and is a joint venture between OPTI Canada Inc. and Nexen to produce a premium synthetic crude. Plants larger than this have been designed on paper, but a break point is believed to occur where economy of scale will no longer be valid. Oxygen purities above 98% are considered to have a significant cost impact on the cryogenics portion of the process; dropping purities below 95% does not create a cost benefit because of the reduction in the main air compressor discharge pressure requirements (20), therefore the optimum falls between 95%–98%.

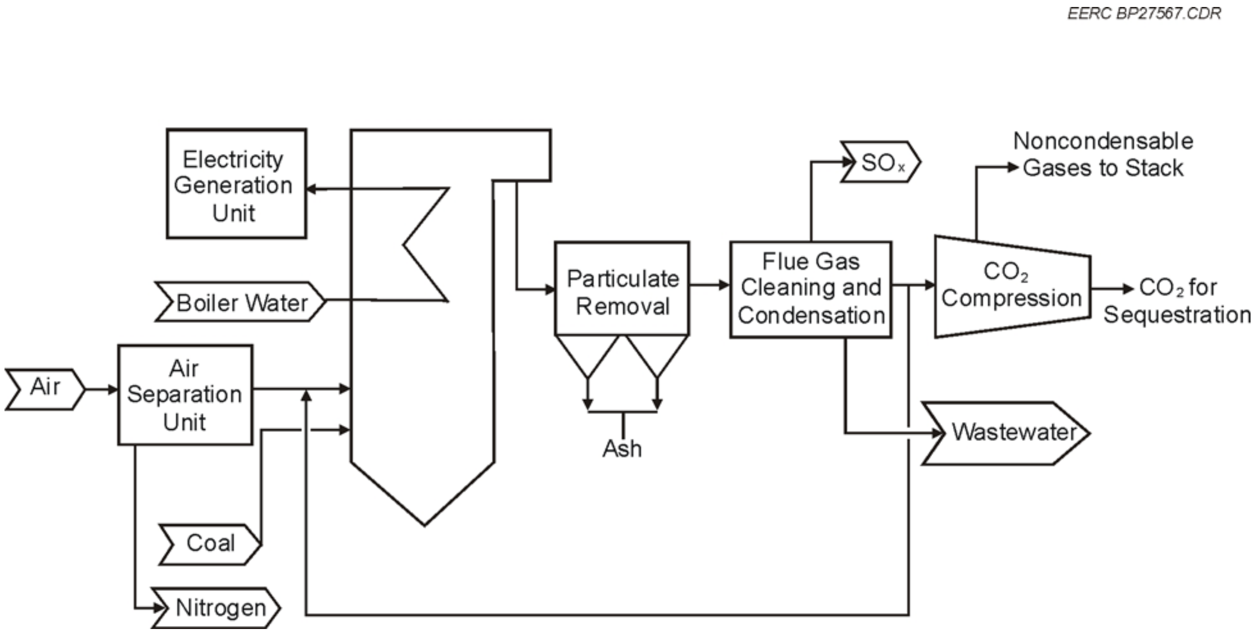

Figure 8. Overall flow diagram of an oxyfuel system for CO_2 capture.

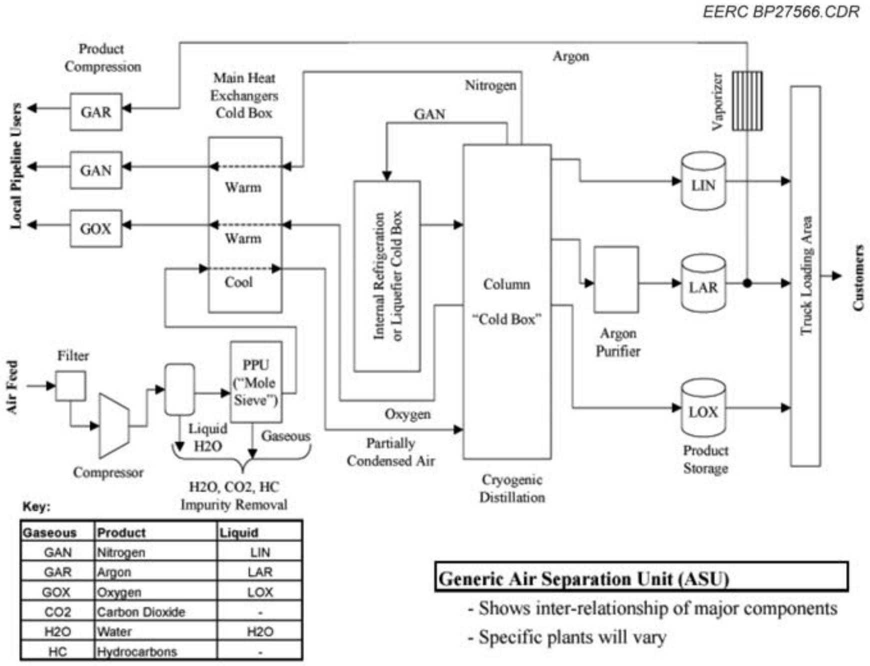

Figure 9. A generic process flow diagram for a cryogenic ASU (courtesy of Universal Industrial Gases Inc.).

Several alternative methods for commercial O_2 production exist, but currently only offer capacities below 100 tpd, and mostly do not produce O_2 at a sufficient purity for use in oxycombustion (28). Current research is ongoing to provide a technology that would separate air more efficiently in order to reduce costs. If a breakthrough technology is discovered, the economics of oxycombustion improves drastically. The alternative technologies that are currently available include the following:

- ***Polymeric membranes:*** This process makes use of selective diffusion of O_2 through membranes. It is suitable for very small flow rates and for O_2 purities below 50%.

- ***Adsorption:*** This process involves selective/regenerative O_2 or N_2 adsorption onto surfaces, typically through pressure swing. Current capacities are less than 200 tpd at purities of 75%–90%.

- ***Chemical:*** This method entails selective and regenerative O_2 absorption into liquid salts; it remains an experimental technology.

- ***Ion transport membranes:*** These membranes use either a platelike (Air Products) or tubular (Air Liquide) perovskite membrane assembled in modules to separate O_2 from hot, pressurized air. The technology is currently restricted to 1 tpd, and it is considered unlikely that it could scale up to oxycombustion scales (29).

A study performed by ALSTOM Power, Inc., determined that for a cryogenic air separation facility producing roughly 9000 tons per day of O_2, the required total plot plan area is 555 by 620 ft. These numbers are supported by information from Air Products (20).

3.2.4 Boiler System Modifications

Several boiler system modifications have been identified by a study done by ALSTOM Power, Inc., and a discussion follows for each.

3.2.4.1 Boiler

The use of oxyfuel as a technique to capture CO_2, impurities in the flue gas that dilute the concentration of CO_2, such as N_2, increases the costs associated with capture. In order to minimize the impurities, the boiler island should be inspected for air leaks into the system and should be sealed to limit infiltration.

3.2.4.2 Ducts, Dampers, and Fans

Several new ducts and dampers will be required; specific needs will be determined on a case-by-case basis. Ductwork will be required for new equipment needed for the process, including oxygen heaters, feedwater heater, booster fan, and the flue gas recycle line.

3.2.4.3 Oxygen Heater

Cold oxidant oxygen from the ASU must be heated. This is accomplished in an oxygen heater that uses hot flue gas to heat the oxygen and recovers sensible heat. This oxidant is then mixed with recycled flue gas before being injected into the boiler.

3.2.5 Effects on Existing Equipment

Flue gas compositions for oxyfuel combustion vary greatly compared to the flue gas produced by typical coal-fired utility systems. Studies have been done to investigate the effect the flue gas composition has on electrostatic precipitators (ESPs) and FGD units. The results are discussed below.

3.2.5.1 Electrostatic Precipitators

Literature reviews indicate that ESP performance degradation is not expected for an ESP operating with rich flue gas. CO_2 has electron affinity resulting in a significant increase in voltage, beyond corona onset voltage, before spark breakdown occurs.

3.2.5.2 Flue Gas Desulfurization

In an FGD unit, SO_2 forms a much stronger acid than does CO_2. It is important to note that the SO_2 is about 35 times more soluble in water than CO_2. The pH of a scrubbing solution with absorbed SO_2 is about 4.5. At this pH level, the CO_2 is basically insoluble in water. It has been predicted that the SO_2 capture efficiency will decrease, although only by a small percentage (20).

3.2.6 Carbon Dioxide Separation and Compression

Once the small quantities of impurities are removed from the flue gas, it is sent to the final stage of the oxyfuel combustion process for CO_2 capture. This phase of the process consists of flue gas cooling, gas compression, gas drying, CO_2 condensation and stripping, and final pumping of the CO_2. The process is quite simple. The CO_2-rich gas is first cooled and compressed to a liquid. The liquid is then passed through a CO_2 stripper to reduce the O_2 and N_2 concentrations to an acceptable level. During the final step, the CO_2 liquid is pumped to a high pressure so that it can be economically transported. A more detailed description follows.

3.2.6.1 Flue Gas Cooling

The flue gas from the existing pollution control devices enters the CO_2 recovery system where it is cooled to around 100°F and is brought to low pressure (essentially subatmospheric) in a gas cooler. At this temperature and pressure, a significant amount of water condenses. Approximately 70%–80% of the gas is recycled back to the boiler, while the remaining one-third is fed to the CO_2 compressor. A direct contact cooler, in which cold water flows countercurrently to the flue gas, is typically used.

3.2.6.2 Flue Gas Compression

The compression system is used to increase the pressure of the flue gas to a pressure high enough to liquefy the stream upon cooling. This is estimated to be around 365 psig. Typical systems used to achieve the desired pressures employ a three-stage, intercooled compression system with intermediate knockout drums to catch the condensed water. Once the flue gas is compressed, it is sent to a flue gas dryer where additional moisture is removed.

3.2.6.3 Flue Gas Drying

It is necessary to dry the CO_2 stream to meet product specifications. Too much water in the CO_2 stream can cause transportation problems, such as pipeline corrosion.

3.2.6.4 CO_2 Condensation and Stripping

From the CO_2 dryer, the gas stream is cooled down further in a CO_2 condenser where, because of the high pressure and low temperature, the CO_2 is liquefied. Once cooled, the remaining impurities are flashed into vapors that contain about 80 wt% of the inlet O_2 and N_2. The liquefied CO_2 is now ready to be pumped to the high pressures necessary for transport. This is achieved by using a heavy-duty pump to bring the pressure to the range of 1600 to 2300 psig.

4.0 MODEL DEVELOPMENT

The scope of the project was to develop a user friendly, Excel workbook-based model of both the amine-scrubbing technology and oxyfuel combustion. The models were to contain mass and energy balances and the economics of retrofitting the necessary equipment to an existing coal-fired power plant. Two separate workbooks were created to estimate the flows and costs associated with retrofitting existing units in North Dakota with the chosen technologies. Different methods were used to develop the models in Excel. General combustion calculations were used to initiate the flow sheets for both models. Once initial flows were generated, performance calculations were developed by using data from literature reviews and IECM developed by Carnegie Mellon University. Guidelines specified by NETL for CO_2 capture feasibility were followed in the creation of the Excel models. The following section describes the development of the two Excel workbook-based models.

4.1 Integrated Environmental Control Model

IECM is a tool for calculating the performance, emissions, and cost of a fossil-fueled power plant. The model was developed by Carnegie Mellon University with support from NETL. The amine-scrubbing portion is one module within the model that can be used to predict the mass, energy, and cost of an amine-based CO_2 removal system. Also included is the more recently developed module for oxycombustion for CO_2 removal. This module also predicts the mass, energy, and cost required for an oxyfuel combustion system. Aspects of these two modules of IECM were used in preparation of the Excel version models prepared for the Lignite Energy Council (LEC).

IECM was built using two process simulators: ASPEN Plus and ProTreat. These two process simulators were used to derive the performance equations used in the model. ProTreat™ is a true mass- and heat-transfer rate-based engineering software tool that was developed by optimized gas treating. It simulates processes used to remove H_2S, CO_2, and mercaptans from a variety gases by absorption into thermally regenerable aqueous solutions containing one or more amines (ProTreat 2002). ASPEN Plus is a widely used process simulator that is primarily used during the design and optimization of steady-state process plants.

4.2 EERC-Developed Amine-Scrubbing Model

The amine-scrubbing model was developed at the request of the LEC. The model predicts the mass flow rates for a coal-fired utility burning a lignite and equipped with an amine-scrubbing system for CO_2 capture. The model also characterizes the performance and costs of the amine-scrubbing system. The following is a discussion of the specific worksheets in the Excel model.

4.2.1 General Mass Balance

The amine-scrubbing model begins with a general mass balance, with several inputs that can be changed depending on the user's process. The major inputs include plant size, capacity factor, gross unit heat rate, and fuel characteristics. Other inputs exist but typically are not changed. Default values are listed for cases where inputs are not known by the user. The general mass balance is a set of combustion equations taken from two text books on combustion (30, 31). This portion of the model estimates the flows of different components in the flue gas, primarily particulates, SO_2, CO_2, N_2, O_2, and Hg. It also computes total flows and temperatures in a variety of units. The general mass balance sheet was primarily developed by the EERC. This basic general mass balance model was then tailored for the specific needs of the overall amine-scrubbing model.

4.2.2 Particulate and SO_x Control

The next worksheets in the model are part of a particulate matter and SO_x control module, which predicts the inlet and outlet flows through a user-specified particulate control and SO_x control technology. The particulate control model relies on the user to select the technology used, specify the collection efficiency, specify the percentage of ash from the boiler, and percentage of SO_2 removal with the ash. The percentage of ash from the boiler is simply the split of bottom ash and the amount that exits the boiler. Sixty-five percent is entered as a default value. The amount of SO_2 removed with the ash is an estimate that may or may not be known for a given system. The default value is 25%. The SO_x control worksheet follows the particulate worksheet. Again, in this worksheet the user specifies the technology used at the given plant and flows are calculated. Inputs included on the SO_x control model include SO_2 capture percentage and ash removal percentage. Equations for these two sheets were developed from equations in combustion text books (30, 31).

4.2.3 Amine Performance Model

The amine performance model is a set of equations designed to characterize an amine-scrubbing system. The responses of this model are used in calculating the operating and capital costs of an amine-scrubbing system. This set of equations was developed through a literature search on the topic as well as IECM. Parameters calculated from the previous worksheets are used in conjunction with input parameters for the CO_2 system. The performance of the CO_2 capture system can be controlled by the input parameters given at the start of the worksheet. A brief description of the important parameters, both input and calculated, follows:

- **Lean sorbent CO_2 loading (π_{lean}):** Upon the application of heat, the CO_2-rich sorbent is regenerated. Ideally, all of the CO_2 is released from the MEA, but in reality not all of it is. The amount of lean sorbent CO_2 loading depends mainly on the initial CO_2 loading in the sorbent and the amount of heat supplied for regeneration. The default value for this is 0.20 based on the range of literature values reported, 0.10–0.25.

- **CO_2 capture efficiency (η_{CO2}):** A literature search revealed that the CO_2 capture efficiency is generally 90% for an amine-based CO_2 capture system; CO_2 capture efficiencies of amine-based systems as high as 96% have been reported. The default value for this model is 90%, but it can be changed to the user's desired level.

- **MEA concentration (C):** The most commonly used MEA technology is supplied by Fluor Daniel. Currently, its technology uses a mixture of MEA with water at a concentration of 30% w/w with the use of corrosion inhibitors. Other suppliers of the technology, who do not use inhibitors, use a lower MEA concentration, typically ranging from 15% to 20% w/w. The default value in the model is 30%, but can be changed at the user's discretion. The acceptable range found in the literature is between 15% and 40%.

- **Acid gas removal efficiency:** The acid gases present in the flue gas (i.e., SO_x, NO, and HCl) are much more reactive towards the MEA than is CO_2. When the gases react with the MEA, they form HSS that cannot be broken down. This causes a permanent loss of MEA sorbent that can be estimated according to the stoichiometry of their reaction with MEA. Removal efficiencies for these gases were determined from literature sources and can be seen in Table 3.

- **MEA loss:** A small quantity of MEA is lost through various unwanted reactions in spite of dilution with water and the use of inhibitors. Long-chained compounds, formed through polymerization reactions and the oxidation reactions forming organic acids and liberating ammonia are the two main unwanted reactions that occur in the system. In general, the loss of MEA can be estimated as 3 lb MEA/ton $CO_{2,}$ with 50% coming from the polymerization reaction and the remaining 50% from the oxidation reaction. Other sources of MEA loss exist in the creation of HSS and NH_3 generation; a reclaimer is used to regenerate some of this loss. The sum of the MEA losses that is reclaimed is used to calculate the amount of makeup MEA necessary to maintain steady-state CO_2 capture.

- **Temperature of the flue gas entering the amine system ($T_{fg\ in}$):** Ideally, the temperature of the flue gas entering the amine system should be 45°–50°C (113°–122°F). This is usually achieved by the installation of a direct-contact cooler.

- **Heat-to-electricity equivalence factor:** The heat-to-electricity equivalence factor is built into the equation that is used to calculate the auxiliary power for the amine system. This factor varies depending on the source of the energy coming from the system. The extraction of low-pressure steam for sorbent regeneration is the major source of energy loss in the system, due to the fact that thermal energy cannot be fully converted into work (Second Law of Thermodynamics). This factor estimates the equivalent loss of power generation capacity because of the heat requirement of the sorbent stripper. Data obtained from literature searches show this factor to lie in the range of 9%–19% for a new plant and 20%–25% for retrofit cases. The value used in this model is 22% for a retrofit case and 14% for a new plant case. The value is automatically changed when the user chooses between a retrofit plant and a new plant in the model. The energy penalty is significantly higher for retrofit cases because most of them have poor heat integration.

- **CO_2 product purity:** The CO_2 product must meet a purity specification in order to be transported, sequestered, or used for other commercial uses. The main impurities to avoid in the final processing stages are N_2 and moisture. Too much N_2 in the stream can cause problems during compression and liquefaction. Excess moisture can cause corrosion in pipelines during transportation and must be controlled. For most applications, the acceptable purity for the CO_2 product is about 99.8%.

Table 3. Removal Efficiencies of Acid Gases in an Amine Absorber

Acid Gas	Removal Efficiency %	MEA Loss, mole MEA/mole acid gas
SO_2	99.5	2
SO_3	99.5	2
NO_2	25	2
NO	0	0
HCl	95	1

- **Sorbent regeneration heat requirement (q_{regen}):** The heat requirement necessary to regenerate the MEA-rich sorbent is expressed as the amount of heat per unit mass of CO_2 captured. Theoretically, the unit of heat required to regenerate the MEA is about 1900 kJ/kg CO_2 (or about 825 Btu/lb CO_2). The actual amount required to regenerate is about 2–3 times higher than the theoretical minimum. Previous studies show a wide range of values for the regeneration heat required. The majority of studies report a heat requirement of 4000 kJ/kg CO_2. In this model, it is a calculated parameter based on the IECM.

- **Enthalpy of regenerating steam (q_{steam}):** Low-pressure steam is used to supply the heat necessary to the MEA stripper. Typically, the low-pressure steam is supplied at 300°C and 60 to 80 psi. Data from the steam tables show the enthalpy of a steam with these parameters to be 2000 kJ/kg steam.

- **Pump and fan efficiencies:** The pump and fan efficiencies used to calculate the auxiliary power necessary were assumed to be 75%.

The remaining parameters that are calculated or input are briefly described by a note in the model (26).

4.2.4 Amine Performance Summary

The amine performance summary worksheet is a summary model that provides the data for the overall performance of the amine CO_2 capture system. In order to calculate the overall effects on the plant, equations to estimate the auxiliary power for other areas in the plant have been used. In this worksheet, the user can select from a list of the type of pollution control technologies that are used for the various pollutants (SO_x, NO_x, particulates, and CO_2). Once the technologies have been chosen, the auxiliary power is estimated and then used to calculate the overall performance of the unit. The major results from this worksheet are the net plant efficiency and net plant heat rate. Other results are calculated, but the majority of the studies use the net plant efficiency for comparison to other technologies.

4.2.5 Amine System Operating and Maintenance (O&M) Costs

The operating and maintenance costs for the amine system include a number of factors and are calculated in the amine O&M worksheet in the model. The worksheet has a number of inputs for which default values have been assigned. The user can change these values if more information is known about the system. A brief description of each cost is discussed below.

4.2.5.1 Variable O&M Costs

- Sorbent: The cost of the sorbent refers to the cost of the MEA makeup requirement in the system. This value is determined by multiplying the amount of makeup MEA required per year as calculated in the performance model by the user input for the cost of MEA. If a cost of MEA is not known, a default value is given.

- Corrosion inhibitor: Inhibitors allow the use of a higher-concentration MEA solution, which increases absorption speeds, while minimizing corrosion. As suggested by IECM, the cost of the inhibitor is estimated to be 20% of the cost of the MEA.

- Activated carbon: The cost of activated carbon is calculated from the physical quantities determined by the performance model.

- Caustic: The cost of caustic is determined from the physical quantities calculated by the performance model.

- Reclaimer waste disposal: The amount of reclaimer waste is calculated in the performance model, and a fee is input in the O&M model by the user or by default.

- Stream (electric equiv.) and electricity: The cost of electricity is the main operating cost. All of the energy costs are assumed to be handled internally by derating the overall power plant based on the calculated auxiliary power requirement. The amine system is charged for the total amount of energy used from the plant because of capture and compression of the CO_2. The cost of electricity is estimated by the model, and the overall cost of electricity per year is calculated.

- Water: Water is primarily required for cooling and process makeup. In comparison, this is a minor expense and is calculated from the performance model and an input cost of water.

- CO_2 transport: The cost of CO_2 transport is an optional cost, and if not chosen, a zero should be entered into the model input for this cost. If it is chosen to be evaluated as part of the CO_2 capture, a default value is given or one can simply be input on a per-ton-of-CO_2 basis.

- CO_2 storage: CO_2 storage costs are also an optional cost. In some cases, this could be viewed as a credit by entering a negative cost in the model. If this is left out of the evaluation, a zero should be entered in the CO_2 storage cost input. If it is included, a default value is given or one can be input on a per-ton-of-CO_2 basis.

4.2.5.2 Fixed O&M Costs

- Operating labor: The operating labor consists of the costs associated with the additional maintenance and labor needed to keep the amine CO_2 capture system running. Default values are given for the inputs necessary to calculate this labor expense and include values for the number of operating jobs, number of operating shifts, and operations labor rate.

- Maintenance material: The maintenance material costs are calculated to be a percentage of the total plant cost (TPC). In the model, a default value of 2.5% of the TPC is entered, but can be changed by the user.

- Administrative and support labor: The administration and support labor is calculated as a percentage of the total labor. A default value of 30% is used, but again can be changed by the user.

All default values are the common values used in the literature.

4.2.6 Amine Cost Summary

The amine cost summary worksheet is where most of the financial calculations are performed. The worksheet starts by calculating the capital costs for the amine CO_2 capture system. The capital cost model is based on the cost per unit flow for the Fluor Daniel system. Equations were developed from IECM to estimate the capital cost per unit flow. From IECM, it is assumed that multiple trains are installed to perform the CO_2 capture operation. The maximum train size has been assumed to be roughly 5500 tons per day. The minimum number of trains required to achieve a desired capture is calculated by IECM. Once this is determined, the cost of equipment is calculated using a chemical engineering price index (32). Prices can be estimated using these indices. Using a scaling factor, an equation can be developed for the capital cost. IECM was used as a guide for the Excel model for development of an equation to determine the capital cost of the necessary equipment. Once the total direct cost is calculated, a set of economic inputs are used to calculate the rest of the economic analysis, resulting in the cost of CO_2 capture in $/ton, the cost of CO_2 avoided ($/ton), and the increase in cost of electricity.

The cost of CO_2 captured in $/ton is calculated by simply taking the total annual cost divided by the total annual CO_2 captured. The cost of CO_2 avoided is the cost of environmental control systems in terms of the cost per ton of pollutant removed or avoided. For an energy-intensive technology such as amine-based CO_2 capture, the cost of CO_2 captured is quite different from the cost of CO_2 avoided. Since the purpose of adding an amine-based CO_2 capture plant is to reduce the CO_2 emissions per net kWh delivered, the cost of CO_2 avoidance is widely used as the economic indicator in this field. Equation 1 shows how this parameter is calculated.

$$\text{Cost of CO}_2\text{ Avoided (\$/ton)} = \frac{(\$/Wh)_{after} - (\$/kWh)_{before}}{(tCO_2/kWh)_{before} - (tCO_2/kWh)_{after}} \qquad \text{[Eq. 1]}$$

4.3 EERC-Developed Oxyfuel Model

The oxyfuel model was developed as requested by the LEC. The model predicts the mass flow rates for a coal-fired utility burning a lignite and equipped with an oxyfuel configuration for CO_2 capture. The model also characterizes the performance and costs of the oxyfuel system. The following is a discussion of the specific worksheets in the EERC-developed Excel model.

4.3.1 General Mass Balance

The oxyfuel model begins with a general mass balance with several inputs that can be changed depending on the user's process. The major inputs include plant size, capacity factor, gross unit heat rate, and fuel characteristics. Other inputs exist, but typically are not changed. Default values are listed for cases where inputs are not known by the user. The general mass balance is a set of combustion equations derived from two text books on combustion and are similar to the ones used in the amine CO_2 capture model (30, 31). This portion of the model estimates the flows of different components in the flue gas, primarily particulates, SO_2, CO_2, N_2, O_2, and Hg. It also computes total flows and temperatures in a variety of units. This basic general mass balance model was then tailored for the specific needs of the overall oxyfuel model. The general mass balance for the oxyfuel model varies from the amine-based system model in that oxygen is calculated in the place of combustion air. Also the calculation for the amount of nitrogen in the system is different, due to the use of almost pure oxygen and not typical combustion air rich in nitrogen. In this model, the inlet and outlet flow of the boiler are calculated including air leaks and the flue gas recycle stream.

4.3.2 Particulate and SO_x Control

The next worksheets in the model are part of a particulate matter and SO_x control module, which predicts the inlet and outlet flows through a user-specified particulate control and SO_x control technology. The particulate control model relies on the user to select the technology used, specify the collection efficiency, specify the percentage of ash from the boiler, and specify the percentage of SO_2 removal with the ash. The percentage of ash from the boiler is simply the split of bottom ash and the amount that exits the boiler. Sixty-five percent is entered as a default value. The amount of SO_2 removed with the ash is an estimate that may or may not be known for a given system. The default value is 25%. The SO_x control worksheet follows the particulate worksheet. Again in this work sheet the user specifies the technology used at the given plant, and flows are calculated. Inputs included on the SO_x control model include SO_2 capture percentage and ash removal percentage. Equations for these two sheets were developed from equations in combustion text books (30, 31).

4.3.3 Oxyfuel Performance and Capital Cost

The oxyfuel performance and capital cost model was developed to characterize and predict the important process parameters necessary to determine both the capital and O&M costs. The model is set up in three main modules that are used to analyze the different process areas of the plant. These areas are the ASU, flue gas recycle and purification, and the overall energy summary of the plant. Also included on this worksheet are the overall flow model for the flue gas recycle and the purification module. In the creation of this model, no attempts were made to quantify changes due to reduced gas flow from oxyfuel combustion. These changes will vary from site to site, making it difficult to incorporate in the model. A brief description of the model follows:

- Air separation unit: The ASU model is a simplistic model that is used to calculate the flows in and out of the unit as well as the composition of the oxidant sent to the boiler for combustion. This information is then used to calculate the auxiliary power requirement and the capital cost of the ASU. Data for this portion of the model came from literature searches and IECM. IECM was used in developing an equation for estimation of both the auxiliary power requirements and capital cost based on the oxygen purity and the oxidant flow rates. Only one major input, oxygen concentration in the oxidant, is necessary for the model to calculate the necessary parameters. As previously discussed, an oxygen concentration of 95% is the suggested purity level and should be used in the model as a default value.

- Flue gas recycle and purification: The flue gas recycle and purification portion of the model is used to determine the flows and energy requirements of this area of the system. The equipment covered in this section includes boiler modifications, flue gas recycle fan, flue gas recycle ducts, oxygen heater, CO_2 purification system, direct-contact cooler, and the CO_2 compression system. Again, IECM was used to develop the equations necessary to estimate the capital costs.

- Energy summary: The oxyfuel energy summary portion of the worksheet provides the data for the overall performance of the oxyfuel CO_2 capture system. In order to calculate the overall effects on the plant, equations to estimate the auxiliary power for other areas in the plant were used. The user can select the type of pollution control technologies that are used for the various pollutants (SO_x, NO_x, particulates, and CO_2) from a list. Once the technologies have been chosen, the auxiliary power is estimated and then used to calculate the overall performance of the unit. The major results of this worksheet are the net plant efficiency and net plant heat rate. Other results are calculated, but the majority of other studies use the net plant efficiency for comparison to other technologies.

4.3.4 Oxyfuel Operating and Maintenance Cost

The operating and maintenance costs for the oxyfuel system include a number of factors and are calculated in the oxyfuel O&M worksheet in the model. The worksheet has a number of inputs for which default values have been assigned. The user can change these values if more information is known about the system. The worksheet is split into two process modules: ASU and CO_2 purification and transport. A brief description of each cost is discussed below.

4.3.4.1 Variable O&M Costs

Air Separation Unit

- Electricity cost: The cost of electricity is the only variable O&M cost for the ASU. This is largely because air is the raw material and it is a free material. All of the energy costs are assumed to be handled internally by derating the overall power plant based on the calculated auxiliary power requirement. The electricity costs are then determined by using the calculated auxiliary power required and multiplying it by the estimated cost of electricity.

CO_2 Purification and Transport

- Electricity cost: The cost of electricity is the largest O&M cost for the CO_2 purification and transport module. The energy costs are assumed to be handled internally by derating the overall power plant based on the calculated auxiliary power requirement. Equipment that requires electricity that is accounted for in this module are the CO_2 purification system, direct-contact cooler, and CO_2 compression system. The electricity costs were then determined by using the calculated auxiliary power required and multiplying it by the estimated cost of electricity.

- Miscellaneous chemicals: Miscellaneous chemicals are used to further purify the CO_2 stream. This is a minor cost and was calculated by the use of IECM.

4.3.4.2 Fixed O&M Costs

- Operating labor: The operating labor consists of the costs associated with the additional maintenance and labor needed to keep the oxyfuel system running. Default values are given for the inputs necessary to calculate this labor expense and include values for the number of operating jobs, number of operating shifts, and operations labor rate.

- Maintenance material: The maintenance material costs are calculated to be a percentage of the TPC. In the model a default value of 2.5% of the TPC is entered, but can be changed by the user.

- Administrative and support labor: The administration and support labor is calculated as a percentage of the total labor. A value of 30% is used as the default, but again can be changed by the user.

4.3.5 Oxyfuel Cost Summary

The total direct capital costs are calculated from the oxyfuel performance and capital cost worksheet. This number is brought into the oxyfuel cost summary worksheet where a set of economic inputs are used to calculate the rest of the economic analysis. The major results calculated in this portion of the model are the cost of CO_2 captured in \$/ton, the cost of CO_2 avoided (\$/ton), and the increase in cost of electricity. Several other results are calculated, but are generally not used when comparing technologies.

The cost of CO_2 captured in \$/ton is calculated by simply taking the total annual cost divided by the total annual CO_2 captured. The cost of CO_2 avoided is the cost of an environmental control system in terms of the cost per ton of pollutant removed or avoided. For an energy-intensive process, such as CO_2 capture, the cost of capture is quite different than the cost of CO_2 avoided. The cost of CO_2 avoidance is widely used when comparing capture technologies. Equation 1 shows how this parameter is calculated.

5.0 PREVIOUS STUDY RESULTS

Several studies have been completed in the past that have determined the effectiveness of amine- and oxyfuel-based CO_2 capture systems, usually on new plants. Below is a brief summary of some of the previous studies.

ALSTOM Power, Inc.:

- The ALSTOM Power, Inc. study, Engineering Feasibility and Economics of CO_2 Capture on an Existing Coal-Fired Power Plant, compared three CO_2 capture technologies for an existing pulverized coal-fired power plant, American Electric Power's Conesville, Ohio, plant Unit No. 5. The unit is a bituminous coal, tangentially fired boiler with a 2400 psig/1000°F/1000°F steam turbine, an ESP, and wet FGD. The net power generated prior to the theoretical addition of the CO_2 capture systems was 433.8 MW_e, with an efficiency of 35% and a CO_2 emission rate of about 2.0 lb/kWh. Three cases for CO_2 capture were studied: an MEA scrubber system, oxyfuel combustion, and an amine mix scrubber.

- In ALSTOM's Concept A (MEA scrubbing), coal is burned conventionally in air. The flue gases pass through a modified FGD system to reach the necessary levels of SO_2 (10 vppm) in the flue gas before entering the MEA scrubber. After the scrubber, the flue gas is cooled in a direct-contact cooler and routed to the MEA system where 96% of the CO_2 is removed, compressed, and liquefied for use or storage. The remaining flue gases leave through the existing stack. Although the boiler performance is the same for Concept A as it is for the base case, 79% of the intermediate-pressure turbine exhaust is extracted for the regeneration of the MEA. The net plant output is 331 MW. This represents a gross plant output reduction of 132 MW over the base case.

- Oxyfuel combustion was used in Concept B, the second case. A cryogenic ASU supplied the roughly 9000 tons/day of 99% pure oxygen to the boiler. The design called for about two-thirds of the flue gas to be recycled back to the boiler. The cryogenic ASU required 96 MW_e of auxiliary power, and the net plant output decreased by 154 MW_e.

- The final case, Concept C, consists of conventionally burned coal (as in Concept A), but the process uses an optimized mixture of MEA and MDEA in a scrubber installed downstream of the FGD. This mixture of amines cannot be made oxygen-resistant; therefore, it is necessary to convert the excess O_2 into CO_2 by combustion with natural gas over a De-Oxy catalyst upstream of the solvent contactor. The solvent regeneration requires about 72% less energy than is required for Concept A. The boiler performance is again identical to the base case, and as was the case for Concept A, there is a major impact to the steam cycle system: 45% of the low-pressure steam is extracted from the existing steam turbine. The new De-Oxy system produces high-pressure steam that is sent to a new turbine, providing about 20% of the energy required by the stripper reboilers. The total output from the modified steam cycle is 431 MW, which represents a gross output reduction of 32 MW, or about 7% of the base case output.

- Table 4 summarizes the results from the ALSTOM Power study. The results indicate that oxyfuel combustion for CO_2 capture is more economically feasible than the MEA scrubbing technology. The cost of CO_2 avoided, as estimated by ALSTOM Power, Inc., for retrofitting an MEA scrubber is \$82/ton CO_2. Their estimation for the oxyfuel combustion CO_2 avoidance cost was lower at \$59/ton CO_2. It is hard to quantify, based on the economics, which of these technologies is truly better because of the error in the estimation method. Most economic studies are only accurate to ± 30% at best and are typically in the range of ± 50%.

Midwest Geological Sequestration Consortium (Illinois Basin Report):

- The Midwest Geological Sequestration Consortium conducted a study entitled Carbon Dioxide Capture and Transportation Options in the Illinois Basin, to review the suitability of various CO_2 capture technologies for large stationary sources. An economic study was conducted by the group to evaluate the energy and economic performance of three new-construction (i.e., greenfield) power generation utilities with CO_2 capture systems. The three plants were 1) pc with postcombustion chemical absorption (MEA), 2) IGCC with precombustion physical absorption (Selexol), and 3) O_2-enriched coal combustion plants. For comparison, a base plant was developed as a conventional pc power plant without CO_2 capture. Three plant sizes were used in the evaluation: 266, 533, and 1054 MW_g. Two fuel types, an Illinois No. 6 coal and a PRB coal, were evaluated. The analysis included the cost of compressing the CO_2 stream to pipeline pressure.

- Results of the study showed that, depending on the plant size and the type of coal burned, CO_2 avoidance cost is between \$47/tonne and \$67/tonne for a pc + MEA plant, between \$31/tonne and \$46/tonne for an oxygen combustion plant, and between \$14/tonne and \$27/tonne for an IGCC + Selexol plant, in 2003 US\$. These costs equate to \$50/ton to \$72/ton for the pc + MEA plant, \$33/ton and \$49/ton for oxyfuel combustion, and \$15/ton and \$29/ton for the IGCC + Selexol plant, in 2005 US\$.

IEA:

- The IEA study involved determining the performance and cost impact of implementing various CO_2 capture systems into different fossil-fueled power systems. Four CO_2 capture technologies were targeted for evaluation: absorption, adsorption, cryogenics, and membranes. The IEA study was split into two phases. Phase One involved developing a baseline plant for each of the plant types to serve as a comparison case. The baseline plants were designed to be a 500-MW_e net output, with a 35-year operating life. Each of the units pc-fired boiler with wet FGD (pc+FGD), natural gas combined cycle (NGCC), IGCC, and oxyfuel combustion) was designed and evaluated by four different research entities. Because of this, it is suggested that the economic analysis be viewed on a relative rather than an absolute-value basis. The fuels used were 0.86% sulfur (as received) Australian bituminous coal or North Sea “Brent” natural gas.

Table 4. Comparison of Results Determined by ALSTOM Power (20)

	Units	Base Plant	Concept A MEA	Concept B Oxy-Fuel	Concept C MEA-MDEA
Fuel Parameters					
Coal Heat Input, HHV	10^6 Btu/hr	4228.7	4228.7	4140	4228.7
Natural Gas Heat Input, HHV	10^6 Btu/hr	–	17.7	11.4	885.9
Total Heat Input, HHV	10^6 Btu/hr	4228.7	4246.4	4151.5	5114.6
Stem Cycle Parameters	kW				
Existing Generator Output	kW	463,478	269,341	463,056	357,196
CO_2 Removal Turbine Output	kW	0	62,081	0	36,343
De-Oxy Turbine Output	kW	0	0	0	37,751
Total Turbine Output	kW	463,478	331,422	463,056	431,290
Total Auxiliary Power	kW	29,700	76,007	189,709	95,317
Net Plant Output	kW	433,778	255,414	273,347	335,973
Overall Plant Performance					
Net Plant Efficiency, HHV	%	35.01	20.53	22.47	22.42
Net Plant Efficiency, LHV	%	36.66	21.5	23.54	23.71
Normalized Efficiency, HHV: Relative to Base Case	%	100	58.64	64.19	64.04
Net Plant Heat Rate, HHV	%	9749	16626	15188	15223
Net Plant Heat Rate, LHV	%	9309	15872	14500	14395
Incremental Cost of Electricity*					
w/o Replacement Power	cents/kWh	N/A	7.4	5.3	10.1
with Replacement Power	cents/kWh	N/A	5.2	4.1	7.9
Cost of CO_2 Avoided					
w/o Replacement Power	$/ton	N/A	82	59	118
with Replacement Power	$/ton	N/A	64	50	98

* Corrected to 2005 US$.

- The pc+FGD plant was designed by the Coal Research Establishment of British Coal Corporation and consisted of a conventional, subcritical pressure steam cycle and a natural circulation boiler, an ESP, and a wet FGD for 90% SO_2 capture. The NGCC plant was designed by the Norwegian Institute of Technology and was configured conventionally, using conventional components (two combustion turbines, two heat recovery system generators, one steam turbine, etc.). The IGCC plant was developed by the Netherlands Research Foundation and incorporated two high-pressure Texaco gasifiers with syngas cooling, two expanders, two combustion turbines, one heat recovery steam generator, one steam turbine, and two elevated-pressure ASUs. Selexol, a cold-gas cleanup method, along with tail gas incineration allowed the plant to operate with 99% sulfur removal. The oxyfuel combustion plant was designed by the Centre for Energy Research in Northern Ireland. The plant consisted of a cryogenic air separator, a pc boiler with low-NO_x burners, an ESP, and an air heater.

- Table 5 shows the results of Phase Two, when the CO_2 capture technologies were added to the plants for all CO_2 capture technologies for the pc + FGD case. In this study, the CO_2 capture technology that was determined to be the most cost-effective was the MEA scrubber technology. The plant efficiency reduced from 40% (LHV) to 29%, and the cost of electricity approximately doubled (in 1992 dollars). The values in the tables were corrected to 2005 dollars for comparison to other studies. It should be remembered that this study was done for a greenfield unit burning a bituminous coal. New units have the advantage of designing for the low-pressure steam extraction required to regenerate the MEA solvent, which can dramatically decrease the energy penalty and costs associated with stripping the MEA. Table 6 shows the results for MEA scrubbing for the different types of plants. Since amine scrubbing is the only technology evaluated during the IEA study that pertains to this study, it is the only technology presented in the table. The cost of CO_2 avoidance for the oxyfuel unit does not relate to this LEC study of North Dakota power plants because it is for a new unit rather than a retrofit. In a new oxyfuel unit, the costs associated with the ASU are not considered part of the CO_2 capture costs, as an ASU is required for plant operation. However, when considering oxyfuel combustion as a retrofit alternative for CO_2 capture, the ASU cost becomes part of the CO_2 capture cost. The ASU represents the majority of the oxy-fuel process expense (33).

Parsons Study:

- The Parsons study was performed by Parsons Energy and Chemical Group, Inc. The project, cofunded by the Electric Power Research Institute and DOE, determined the performance and the cost of applying amine-scrubbing CO_2 capture technology to three greenfield power generation systems: pc+FGD, NGCC, and IGCC. The pc and NGCC cases employed an amine system in the typical postcombustion configuration. The IGCC case employed a physical absorption system because of the elevated pressures of the syngas. The plants were designed to burn either an Illinois No. 6 coal (\$1.26/MMBtu) or natural gas (\$2.7/MMBtu). The CO_2 product stream was compressed to 1200 psig and required a 40°F dew point, a 1.25% H_2 limit, and sulfur limits of 100 and 50 ppm for SO_2 and H_2S, respectively.

- The pc+FGD plant was designed to be conventional in configuration, consisting of a pc-fired boiler, low-NO_x burners, air staging, SCR, an ESP, and a wet FGD. For CO_2 capture, an inhibited aqueous MEA-scrubbing technology was used for chemical absorption. The NGCC plant was evaluated at two different plant configurations. The first unit consisted of two General Electric (GE) combustion turbines, each with its own HRSG that supplied steam to a single steam turbine. The second case used a single GE combustion turbine with one heat recovery steam turbine and one steam turbine. The IGCC plant consisted of an E-Gas^{TM}-type gasification technology, porous metal candle filter for particulates, and a fire tube-type boiler for syngas cooling. The syngas travels through two absorption stages, the first to remove H_2S and the second to capture CO_2.

- The results of the Parsons study can be seen in Table 7. All costs in Table 7 are in 2000 US\$ except for the cost of CO_2 avoided, which was corrected to 2005 US\$ for comparison. The study shows that for a supercritical pc boiler burning an Illinois No. 6 bituminous coal and employing an amine scrubber for CO_2 removal, the CO_2 avoided cost is \$58/ton.

Table 5. Feasibility of CO_2 Capture Technologies for a pc-Fired Power Plant (IEA study)

Unit	Plant Efficiency, % (LHV basis)	Power Cost, mills/kWh	CO_2 Avoided Cost, \$/ton[1]	CO_2 Emission Rate, g/kWh
Baseline Plant	40	65		829
Plant with CO_2 Capture by:				
Absorption	29	98	\$42	116
Adsorption w/PSA	28	151	\$101	57
Adsorption w/TSA	29	236	\$317	335
Membrane	31	110	\$56	194
Membrane w/MEA	30	107	\$54	222

[1] 2005 US$.

Table 6. Feasibility of CO_2 Absorption on Alternative Power-Generating Technologies (IEA study)

	PC+FGD	NGCC	IGCC	pc with O_2/CO_2
Baseline Plant				
Plant Efficiency, % LHV	40	52	42	33
Specific Plant Cost, \$/kW	1397	927	2061	2699
COE, mills/kWh	65	46	70	103
CO_2				
kg/sec/MWh	0.23	0.11	0.21	0.28
mol% wet	13	3	7	63
mol% dry	14	4	7	91
Plant with Absorption				
CO_2 Capture Efficiency, %	90	85	82	99
Plant Efficiency, % LHV	29	42	36	30
Specific Plant Cost, \$/kW	2432	1805	3169	4096
COE, mills/kWh	98	70	83	124
CO_2 Avoided Cost, \$/ton	42	66	27	19

* 2005 US$.

- Bituminous coals offer some advantages over lignite in terms of CO_2 removal in that they are typically lower-moisture and higher-heating-value coals. These characteristics produce less CO_2 on a heating-value basis, reducing the size of equipment and creating lower operating expenses (33).

American Air Liquide:

- Air Liquide's study involved determining the impact of 90% CO_2 capture on a new pc plant using two technologies, amine (MEA) scrubbing and oxyfuel combustion. The CO_2 capture system did not include the costs associated with CO_2 cooling, drying, and compression to pipeline pressure. The base plant was designed to be 501 MWe, with an efficiency of 37% HHV basis. The plant was fueled by a PRB subbituminous coal and included low-NO_x burners with SCR, a limestone spray dryer, an AC system for Hg control, and an ESP. The first CO_2 capture technology evaluated was an MEA scrubber system that captured 90% of the CO_2. The system was installed in the typical postcombustion configuration downstream of the ESP. In this case, the net plant output decreased to 388 MWe, and the net plant efficiency decreased to 28.6%. The second technology was oxyfuel combustion and consisted of the typical oxyfuel system (ASU, recycled flue gas). With the air N_2 eliminated, several cost advantages were realized, including a reduction of flue gas treatment costs and elimination of the SCR. Despite the ASU requiring 100 MWe of auxiliary power, a higher net output is achieved at 405 MWe.

- Table 8 summarizes the energy efficiency of the plant both with and without CO_2 capture technologies. This study determined that, despite the high cost of the ASU, oxyfuel combustion imposes smaller energy efficiency and cost penalties on the plant than MEA scrubbing. The cost of CO_2 avoided for the MEA case was $50/ton, whereas the oxyfuel combustion case was less than half, at $21/ton. The economic year is unknown and was assumed to be 2004 (the previous year of the published year of the report) and corrected to 2005 US$. It should be noted that these costs do not include CO_2 cooling, drying, and compression to pipeline pressure, which can significantly impact the expenses associated with CO_2 capture for sequestration. Figures 10 and 11 summarize the economics of the study (33).

Table 7. Comparison of Results from the Parsons Study

Plant Type:	NGCC	IGCC	SC–pc	USC–pc
Parsons Case No.:	1D	3B	7C	7D
Without CO_2 Capture				
Fuel	Natural Gas	Illinois No. 6	Illinois No. 6	Illinois No. 6
Fuel Cost, $/MMBtu (HHV)	2.7	1.26	1.26	1.26
Plant Net Output, MWe	384.4	424.5	462.1	506.2
Plant Efficiency, % HHV	53.6	43.1	40.5	42.7
Total Plant Cost, $/kW	496	1263	1143	1161
COE, mills/kWh**	30.7	45.1	45	44.3
CO_2 Emissions				
lb/kWh	0.745	1.582	1.707	1.618
kg/kWh	0.338	0.718	0.774	0.734
With CO_2 Capture*				
Parsons Case No.	1B	3A	7A	7B
Fuel Cost, $/MMBtu (HHV)	2.7	1.26	1.26	1.26
Plant Net Output, MWe	310.8	403.5	329.3	367.4
Plant Efficiency, % HHV	43.3	37	28.9	31
Total Plant Cost, $/kW***	943	1642	1981	1943
COE, mills/kWh**	48.8	56.4	74.4	71.6
CO_2 Emissions				
lb/kWh	0.088	0.162	0.238	0.222
kg/kWh	0.04	0.073	0.108	0.101
CO_2 Avoided Cost, $/ton****	80	23	58	57

* Plants include CO_2 processing and compression to 1200 psig pipeline pressure.
** 20-year levelized value with 80% capacity factor.
*** Year 2000 US$.
**** Year 2005 US$.

Table 8. Summary of Plant Performance for the Air Liquide Study

Boiler Oxidant	Air	Air	Oxygen	Oxygen
Flue Gas Recycle	No	No	Yes – Undried	Yes – Dried
CO_2 Concentrated*	No	Yes	Yes	Yes
CO_2 Technology	None	MEA scrubbing	Oxyfuel	Oxyfuel
Steam Turbine Power, MWe	533	434	533	533
Plant Auxilary Power, MWe				
ASU	N/A	N/A	100	104
Other	31	47	24	24
Total	31	47	124	128
Net Plant Power, MWe	501	388	408	405
Net Plant Efficiency, % HHV	37	28.6	31.4	29.9

* Plant does not include CO_2 processing or compression for transport.

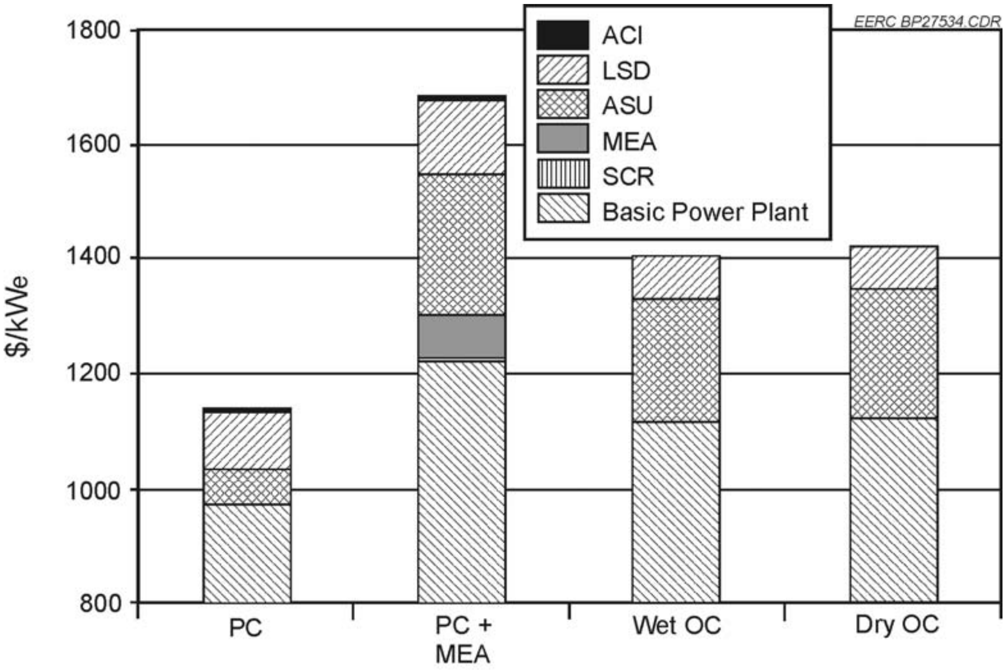

Figure 10. Capital cost breakdown for plants considered during the Air Liquide study (33).

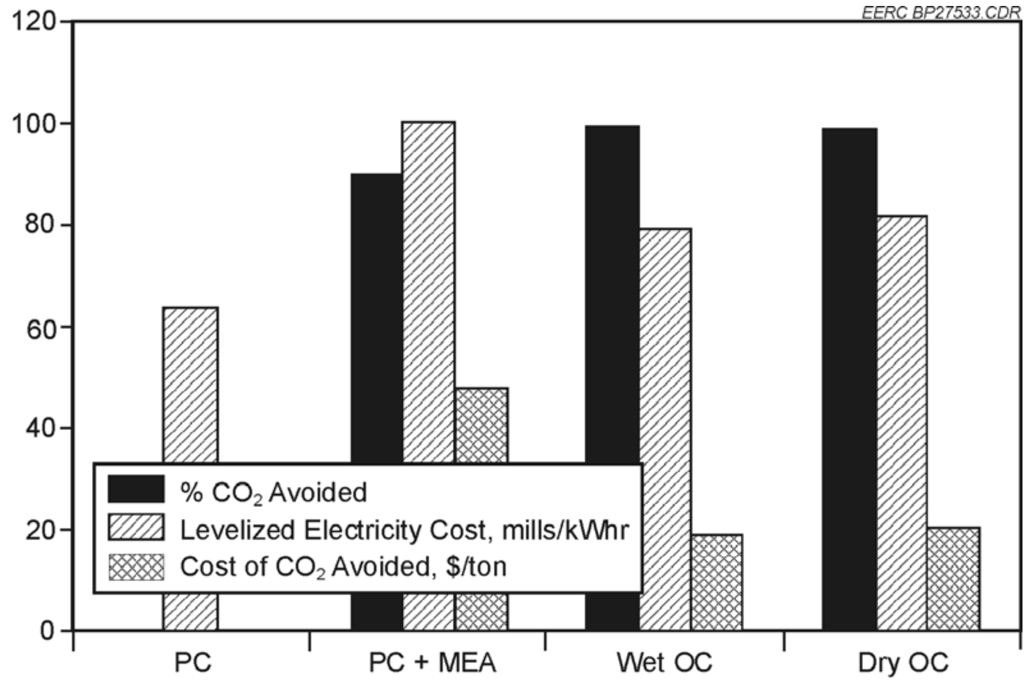

Figure 11. Economic analysis for the Air Liquide study showing the cost of electricity and the cost of CO_2 avoided (33).

6.0 MODEL RESULTS

A case study was run on each of the Excel models, the amine-scrubbing and oxyfuel combustion models, to compare to the study results that have just been summarized. For the base plant, a 500-MWg pc system was used. The unit was equipped with low-NO_x burners, an ESP, and a wet FGD. A North Dakota lignite was chosen as the fuel, with the lignite analysis specifically chosen by NETL's CO_2 capture system guidelines. The coal is from the Beulah–Zap coal seam located in Mercer County, North Dakota; its properties can be seen in Table 9. A capacity factor of 75% was used along with a gross unit heat rate of 9550 Btu/kWh. The ESP is set up to remove 99.9% of the particulate matter. It is assumed that only 65% of the ash from the boiler enters the ESP and the remaining 35% exits in the bottom ash. The wet FGD is assumed to remove 98% of the SO_2 entering the system.

For the amine-scrubbing case, when the coal is combusted, the flue gas produced contains 13.7 mol% CO_2, typical for a lignite. Table 10 displays the calculated flue gas components from the combustion of this coal. The CO_2 capture system is set to remove 90% of the CO_2 in the flue gas, using a 30 wt% MEA solution in the absorber column. The CO_2 product is then compressed and pumped to a final pressure of 2000 psig for transportation. The installation of the MEA scrubber is assumed to be on a retrofit case, which increases the amine scrubber power requirement because of the inefficient extraction of steam. Table 11 shows a summary of the performance of the unit equipped with and without the amine-scrubbing system. The net plant efficiency without CO_2 capture is roughly 33%, which reduces to about 20% when the system is equipped with the amine system. The net output of the plant with CO_2 capture is 282 MW, with a net heat rate of 16,950 Btu/kWh, or approximately 6500 Btu/kWh higher than the base case without CO_2 capture.

The O&M costs were calculated using the model and can be seen in Table 12. The largest O&M costs are from the steam on an electric-equivalence basis, and the makeup sorbent required to sustain 90% capture of CO_2. The total capital cost is calculated to be about $120 million, and a 1.1 retrofit difficulty factor is applied to bring the total direct capital cost to $129 million. Table 13 shows the major results of the amine-scrubbing model. For this system, the cost of CO_2 avoided is $64.30/ton of CO_2. The increased cost of electricity is $62.17/MWh, bringing the total cost of electricity to 10.8 cents/kWh. The results of the model correspond well to the results of the other studies.

The Excel oxyfuel model was evaluated under the same fuel conditions. The base plant is a 500-MWg unit, with a net heat rate of 9025 Btu/kWh and a net plant efficiency of about 33%. The net heat rate is slightly better in the oxyfuel case because the boiler efficiency increases as a result of the use of an oxygen-enriched oxidant for combustion. The ESP is set up to remove 99.9% of the particulate matter. It is assumed that only 65% of the ash from the boiler enters the ESP and the remaining 35% exits in the bottom ash. The wet FGD portion of the model can be used to calculate the allowable removal necessary for the system; in this case, it is 68%.

When the coal is combusted in the oxyfuel case, it produces a flue gas with a CO_2 concentration of 50.2 mol%. After the recycle and air leak considerations, the final CO_2 concentration is 60 mol% (75 wt%). Table 14 shows the flue gas composition as calculated by the model for the Beulah–Zap coal. The total flue gas flow rate is approximately one-third less than that of the amine-scrubbing case. Only small advantages are seen since this is a retrofit case and the existing equipment is already designed to handle the larger flow rates of flue gas. If the plant were to be built as a greenfield plant, greater advantages could be seen in NO_x and smaller pollution control equipment sizes. Table 15 provides the performance summary of the oxyfuel system, which shows that the oxyfuel system is more efficient than the amine-scrubbing unit. The net plant efficiency is 25%, and the net plant heat rate is 13,777 Btu/kWh for the oxyfuel system.

The O&M costs of the oxyfuel system are shown in Table 18. The largest expenses are the electricity for both the ASU and the purification and compression portions of the process. A fuel credit was taken in the O&M of this plant because of the reduced amount of coal needed since the boiler efficiency increased. The total O&M costs are $58 million/year. The total direct capital cost for the oxyfuel retrofit is $324 million. The ASU makes up $230 million of the total capital expense. As discussed earlier, the capital cost of the ASU was calculated based on the costs from the IECM. This cost of $230 million is on the high side for what is reported in previous studies ($150–$180 million [2005 US$]). Table 17 shows the major economic results of the model. The cost of CO_2 avoided is $63.87/ton CO_2, with the increased cost of electricity being $62.48/MWh. The increased cost of electricity brings the total cost of electricity to 10.9 cents/kWh. The results of the study agree well with the results of previous studies. The ALSTOM Power Inc. study reported a cost of CO_2 avoided of $59/ton CO_2 for its oxyfuel system, which is within error limits of the values calculated by this model.

Table 9. Fuel Characteristics as Displayed in the Amine-Scrubbing Model

Fuel Characteristics (as-received)		
Coal Name (mine, seam, etc.)	Beulah–Zap	Input
Coal Type (lignite, subbituminous, bituminous, etc.)	Lignite	Input
Heating Value, Btu/lb (as-received basis) HHV	7454	Input
Estimated Based on Dulong's Formula, Btu/lb	7386	Calc
Estimated Based on Revised Mason Formula, Btu/lb	8390	Calc
Ultimate Analysis, % as-received		
Carbon, %	44.6	Input
Hydrogen, %	2.95	Input
Oxygen in Fuel, %	12.32	Input
Nitrogen, %	0.70	Input
Sulfur, %	0.54	Input
Moisture, %	32.24	Input
Ash, %	6.59	Input
Total (should = 100%)	99.96	
Mercury Concentration, dry ppm in coal	0.08	Input
Mercury Concentration, ppm in as-received coal	0.054	Calc

Table 10. Total Flue Gas Composition and Flows from Combustion of Lignite Coal from Beulah–Zap Coal Seam

Major Flue Gas Components	Flue Gas In, tons/hr	MW	lbmol/hr	Kmol/hr	Mol%	Mass%
Nitrogen (N_2)	1668	28.013	119,081	54,014	68.5	66.1
Oxygen (O_2)	83.6	31.999	5224	2370	3.0	3.3
Carbon Dioxide (CO_2)	523	44.01	23,771	10,782	13.7	20.7
Sulfur Dioxide (SO_2)	3.29	64.065	103	47	0.1	0.1
Water	218	18.015	24,193	10,974	13.9	8.6
Argon	27.8	39.948	1392	632	0.8	1.1
Total Wet Flue Gas	2524		173,763	78,818	100.0%	100.0%
Particulate	21.11					
Total	2545					
Total, acfm	1,804,934					

7.0 RETROFITTING NORTH DAKOTA COAL-FIRED UTILITY FLEET

North Dakota lignite-fired systems have high levels of carbon content per unit of energy input which translates into relatively high CO_2 emission levels as compared to other fossil fuel-based power systems. For this reason, it is particularly important for North Dakota lignite users to consider the issues associated with retrofitting the existing fleet with CO_2 capture technology. Calculations have been completed to determine the estimated costs of these changes. In addition, several discussions have taken place with industry representatives to determine what the key issues from their perspective are to make North Dakota's coal-fired utility fleet CO_2 capture-ready. As part of these conversations, several concerns/challenges were identified. Some of these challenges are determining a technology, size and space requirements, and determining where to put the CO_2 once it has been captured. The EERC has prepared several carbon management plans (CMP) to determine what is required to construct a grassroots facility that is CO_2 capture-ready. A discussion of the CMP procedure follows.

7.1 Carbon Management Plans

Because CO_2 emissions receive ever-increasing scrutiny and are the subject of potential regulation in the future, the development of a CMP is critical for any new facility and could be applied when considering retrofit options. A thorough CMP should:

- Describe the facility, the capture technology that has been selected (or the technologies under consideration), and quantify the expected CO_2 emissions.

- Identify and summarize the capacity of geologic sequestration options in the vicinity, including both potentially profitable opportunities such as for EOR, ECBM, or straight sequestration into depleted oil fields or saline formations.

- Identify any infrastructure (e.g., pipelines, booster stations) needed to transport the CO_2 to the geologic sink(s).

- Quantify the CO_2 produced by other sources in the area.

- Discuss the costs associated with capture, dehydration, compression, and transportation of CO_2.

- Identify any environmental risks and discuss mitigation options.

- Identify local terrestrial sequestration opportunities and provide rough cost estimates.

- Include a discussion of emerging carbon markets.

- Offer a prioritized list of suggested carbon sequestration options that appear to be the most appropriate for the new facility and estimate the cost of each.

7.2 Costs Associated with Retrofitting North Dakota Coal-Fired Generating Fleet

North Dakota has several lignite-fired power generating stations which produce a large amount of CO_2. In order to determine the cost of retrofitting the state of North Dakota coal-fired utilities, the IECM was used to calculate the costs and power requirements associated with adding an amine scrubber system to the postcombustion side of all North Dakota's utilities. The IECM was chosen in this analysis because it has been reviewed by DOE. A list of the major coal-fired utilities and a description of their existing pollution control equipment is in Table 18. As seen in Table 18, seven major plants consisting of thirteen units were included in the analysis. Figure 12 shows a map of North Dakota and the plant locations in the state. The majority of these units use a dry scrubber or no control for SO_2. In this case, a wet FGD was added into the cost associated with the addition of an amine-scrubbing system. This was done to reduce the O&M costs associated with solvent use due to the high usage of MEA in the presence of SO_2.

A CO_2 reduction of 60% was chosen as a basis and was based on information seen in the USDOE NETL *Carbon Sequestration Technology Roadmap and Program Plan 2006.* The results indicate a large annual levelized cost of $1.12 billion/yr with an average CO_2 capture cost of $82.60 per ton. This cost includes the capital and O&M costs for the installation of an amine system at each plant as well as the capital and O&M costs to install a wet FGD where necessary. An additional cost would also be added to this number for the costs associated with replacing the lost power production in the state. By capturing 60% of the CO_2 with an amine system, the auxiliary power load increases by 1065 MW, which would then need to be replaced. Several methods to replace this power could be used. Power could be bought off the grid by importing power if it were available. To replace this amount of electricity, it would cost ~ $700 million/yr assuming a charge of $100/MWh, which is an average cost of electricity when considering CO_2 capture with an amine system. Figure 13 shows the sensitivity of the cost of replacement electricity as a function of the cost of purchasing electricity off the grid. Figure 13 shows this cost to vary anywhere from $400 million/yr to $1 billion/yr in the most probable region of electricity purchase price.

Table 11. Performance Summary Calculated from the Amine-Scrubbing Model

Energy Requirements (auxiliary power)		Unit	Direction
Boiler	29.25	MW	Calc.
Pollution Control			
Particulates			
C-ESP	1.177	MW	Calc.
SO_x			
Wet FGD	15.9	MW	Calc.
CO_2			
Amine System	172	MW	Calc.
NO_x			
Low-NO_x Burner (LNB)	0	MW	Calc.
Total Auxiliary Power Req'd.	218.4	MW	Calc.
Energy Balance Summary			
Gross Electrical Output	500	MW	Calc.
Net Electrical Output	281.6	MW	Calc.
Primary Fuel Energy In	4775	Mbtu/hr	Calc.
Gross Plant Heat Rate, HHV	9550	Btu/kWh	Calc.
Net Plant Heat Rate, HHV	16,954	Btu/kWh	Calc.
Net Plant Heat Rate Base Plant, HHV	10,939	Btu/kWh	Calc.
Annual Operating Hours	6575	Hrs/yr	Calc.
Annual Power Generation	1.85	BkWh/yr	Calc.
Annual Power Generation, w/o CO_2	2.98	BkWh/yr	Calc.
Net Plant Efficiency, HHV	20.1	%	Calc.
Net Plant Efficiency w/o Amine, HHV	32.8	%	Calc.
Efficiency Difference	12.7	%	Calc.

Table 12. Breakdown of the Operating and Maintenance Cost as Calculated by the Amine-Scrubbing Model (cost in million $/year)

Variable Costs			
Sorbent	M$/yr	$8.2	Calc.
Corrosion Inhibitor	M$/yr	$1.6	Calc.
Activated Carbon	M$/yr	$0.32	Calc.
Caustic (NaOH)	M$/yr	$0.26	Calc.
Reclaimer Waste Disposal	M$/yr	$1.7	Calc.
Steam (elec. equiv.) and Electricity	M$/yr	$52.4	Calc.
Water	M$/yr	4.89E-01	Calc.
CO_2 Transport	M$/yr	$0.0	Calc.
CO_2 Storage	M$/yr	$0.0	Calc.
Total Variable Costs	M$/yr	$65.0	Calc.
Fixed Costs			
Operating Labor	M$/yr	$2.1	Calc.
Maintenance Material	M$/yr	$7.0	Calc.
Admin. and Support Labor	M$/yr	$0.62	Calc.
Total Fixed Costs	M$/yr	$9.7	Calc.
Total O&M	M$/yr	$74.7	Calc.

Table 13. Major Economic Analysis Results for the Amine-Based CO_2 Capture System

Levelized Annual Costs (2005 US$)	
Levelized Fixed Charges	$40,431,759
Levelized Operating Costs	$74,683,199
Total Annual Cost, $	$115,114,958
Total Annual Cost, $/kW	$230.23
$/ton CO_2 Captured	$37.20
Increased Cost of Electricity COE, $/MWh	$62.17
Cost of CO_2 Avoided, $/ton	$64.30
Total Cost of Electricity, $/kWh	$0.108

Table 14. Flue Gas Components and Flow Rates of the Oxyfuel System as Calculated

Flue Gas out with Recycle + Air Leak						
Major Flue Gas Components	Flue Gas In/tons/hr	MW	lbmol/hr	Kmol/hr	Mol%	Particulate-Free, wt%
Nitrogen (N_2)	77.74	28.0	5550	2517	5.7	4.5
Oxygen (O_2)	27.98	32.0	1749	793	1.8	1.6
Carbon Dioxide (CO_2)	1285.35	44.0	58,412	26,495	59.8	75.0
Sulfur Dioxide (SO_2)	4.65	64.1	145	66	0.1	0.3
Water	260.40	18.0	28,910	13,113	29.6	15.2
Nitric Oxide	0.66	30.0	44	20	0.0	0.0
Argon	56.49	39.9	2828	1283	2.9	3.3
Total Wet Flue Gas	1713		97,638	44,288	100%	100%
Particulate	19.95					
Total	1733					

Table 15. Performance Summary Calculated from the Oxyfuel Model

Energy Requirements (auxiliary power)		Unit	Direction
Boiler	29.25	MW	Calc.
Pollution Control			
Particulates			
C-ESP	1.03	MW	Calc.
SO_x			
Wet FGD	4.93	MW	Calc.
CO_2			
Oxyfuel	57.4	MW	Calc.
NO_x			
LNB	0	MW	Calc.
Required for Oxyfuel			
ASU	79.9	MW	Calc.
Total Auxiliary Power Req'd.	172.5	MW	Calc.
Energy Balance Summary			Calc.
Gross Electrical Output	500	MW	Calc.
Net Electrical Output	327.5	MW	Calc.
Primary Fuel Energy In	4512.5	Mbtu/hr	Calc.
Gross Plant Heat Rate, HHV	9025	Btu/kWh	Calc.
Net Plant Heat Rate, HHV	13,777	Btu/kWh	Calc.
Net Plant Heat Rate Base Plant, HHV	10,393	Btu/kWh	Calc.
Annual Operating Hours	6575	hr/yr	Calc.
Annual Power Generation	2.15	BkWh/yr	Calc.
Annual Power Generation w/o CO_2	2.98	BkWh/yr	Calc.
Net Plant Efficiency, HHV	24.8%	%	Calc.
Net Plant Efficiency w/o CO_2 Capture, HHV	32.8%	%	Calc.
Efficiency Difference	8.1%	%	Calc.

Table 16. Breakdown of the O&M Costs as Calculated by the Oxyfuel Model (cost in million $/year)

Air Separation Unit		Units	
Electricity Cost, M$/yr	24.31	M$/yr	Calc.
Total Variable	24.31	M$/yr	Calc.
Labor	5.21	M$/yr	Calc.
Maintenance Material	4.59	M$/yr	Calc.
Admin. and Support	1.56	M$/yr	Calc.
Total O&M, M$/yr	35.67	M$/yr	Calc.
CO_2 Purification and Transport		Units	
Misc. Chemicals	0.73	M$/yr	Calc.
CO_2 Transport	0.00	M$/yr	Calc.
CO_2 Storage	0.00	M$/yr	Calc.
Electricity	16.59	M$/yr	Calc.
Fuel Credit	-1.74	M$/yr	Calc.
Total Variable	15.59	M$/yr	Calc.
Labor	2.34	M$/yr	Calc.
Maintenance Material	3.77	M$/yr	Calc.
Admin. and Support	0.70	M$/yr	Calc.
Total O&M	22.40	M$/yr	Calc.
Overall Total O&M	58.07	M$/yr	Calc.

Table 17. Major Economic Analysis Results for the Oxyfuel-Based CO_2 Capture System

Levelized Annual Costs (2005 US$)	
Levelized Fixed Charges	$77,843,171
Levelized Operating Costs	$58,071,378
Total Annual Cost, $	$135,914,549
Total Annual Cost, $/kW	$271.83
$/ton CO_2 Captured	$44.78
Increased Cost of Electricity (COE), $/MWh	$63.12
Cost of CO_2 Avoided ($/ton)	$63.87
Total Cost of Electricity $/kWh	$0.109

New plant construction was also used as a basis for replacement power. The IECM was again used to determine the cost to build a new plant that included an amine system for 60% CO_2 capture. In order to meet the energy shortage created by the installation of the amine systems, an approximate 1600-MW$_g$ plant would need to be built. Current technology for a new lignite-fired facility would most likely be a supercritical or an ultrasupercritical plant and not IGCC because of technical issues with current gasifiers. If one 1600-MW$_g$ supercritical unit is able to be constructed, it would cost ~$2.3 billion in capital with a total levelized annual cost of $577 million/yr. If this unit is not feasible because of size and two 800-MW$_g$ units are needed, the total capital required is $2.6 billion with a total levelized annual cost of $726 million/yr. The

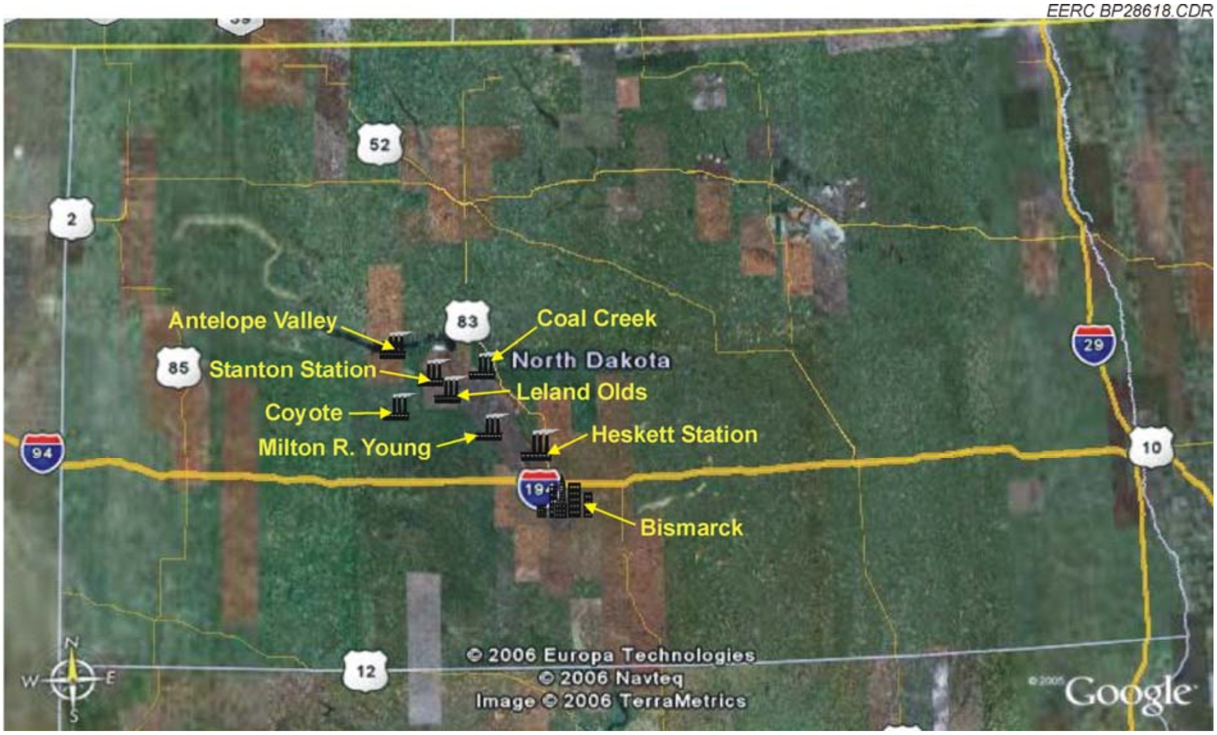

Figure 12. Map of North Dakota's coal-fired utilities.

IECM estimated that for similar ultrasupercritical units, the levelized annual costs would be reduced slightly. Table 19 summarizes the results for the costs associated with capturing 60% of the CO_2 produced by the existing coal-fired power plants in North Dakota. If an annual levelized cost of $700 million/yr were used for the replacement energy costs, the total yearly expense would be $1.82 billion.

8.0 SUMMARY

Several options could be explored to capture CO_2 from lignite-fired power plants, but few of these options exist on a commercial level. The most proven of the commercially available options is amine scrubbing. Although these options exist today, none of them can provide an inexpensive means by which to capture CO_2 in terms of energy or economics. The technologies currently available all require a large amount of energy which, in most cases, will more than double the auxiliary power requirements of the plant. High costs accompany the large energy requirements and would require significant new construction just to maintain the current levels of power to the grid. Research is ongoing to develop ways to provide an economical CO_2 capture technology, with several promising technologies in the research and development phases today. The analysis performed in this study did not take carbon credits into consideration. If a carbon incentive were offered, the economics for each process would improve.

Table 18. Description of North Dakota Plants

						SO$_2$ Control		PM Control	
Plant Owner	**Plant/Unit**	**City**	**MW$_e$**	**Boiler Type**	**NO$_x$ Control**	**Dry FGD**	**Wet FGD**	**Cold ESP**	**Fabric Filter**
Basin	Leland Olds Unit 1	Stanton	216	Wall-fired	LNB/OFA			X	
Basin	Leland Olds Unit 2	Stanton	440	Cyclone	None			X	
Basin	Antelope Valley 1	Beulah	435	T-fired	LNOVF[1]	X			X
Basin	Antelope Valley 2	Beulah	435	T-fired	LNOVF	X			X
Great River	Coal Creek 1	Underwood	540	T-fired	LCN3[2]		X	X	
Great River	Coal Creek 2	Underwood	540	T-fired	LCN3		X	X	
Great River	Stanton Station	Stanton	140	Wall-fired	LNB			X	
Great River	Stanton Station 10	Stanton	60	T-fired	LNB	X			X
MDU	Heskett Unit 1	Mandan	25	Stoker	None			X	
MDU	Heskett Unit 2	Mandan	85	Fluid bed	None	FBC		X	
Minnkota	Milton R. Young 1	Center	235	Cyclone	None			X	
Minnkota	Milton R. Young 2	Center	440	Cyclone	None		X	X	
Otter Tail	Coyote	Beulah	460	Cyclone	None	X			X

[1] Low-NO$_x$ burner and overfire air.

[2] Low-NO$_x$ coal and overfire air (Option 3).

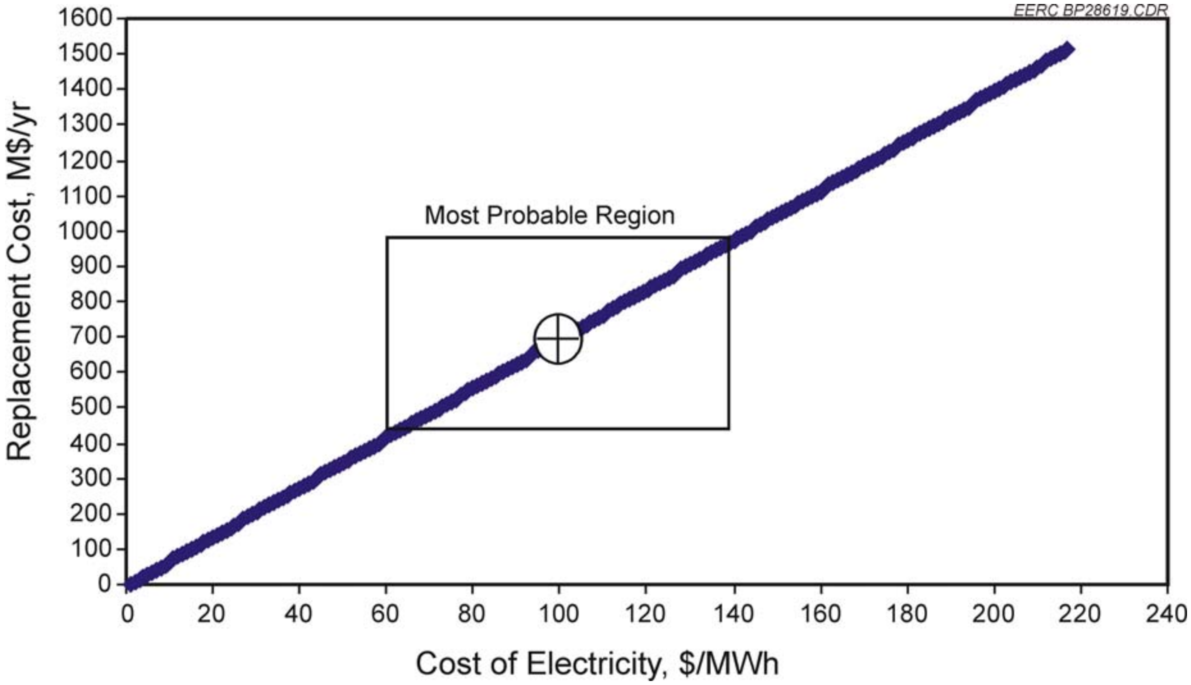

Figure 13. Sensitivity of the cost of replacement electricity as a function of the cost of purchasing electricity off the grid.

A comparison of the two technologies discussed in this report can be seen in Tables 20 and 21. The oxyfuel combustion technology is more efficient, but when costs are compared, the two technologies are statistically the same in terms of cost/ton of CO_2 removed and $/kWh. If the cost for an ASU can be decreased, the oxyfuel process becomes more economically attractive as this is the majority of the costs for the process.

Table 19. Summary Results for the Costs Associated with Capturing 60% of the CO_2 Produced by North Dakota's Current Coal-Fired Utilities by Retrofitting Them with an Amine System

Plant Name	**Antelope Valley**		**Coal Creek**		**Coyote**		**Leland Olds**	
Gross Electrical Output (MWg)	870		1080		460		656	
Amine Scrubber Use (MW)	211		262		112		159	
Wet Scrubber Add./Unit	Yes/Units 1 and 2		No		Yes		Yes/Units 1 and 2	
Wet FGD Use (MW)	28.9		N/A		14.8		21.3	
Net Electrical Output (MW)	575		716		305		435	
CO_2 Captured, Mtons/yr	3.53		4.38		1.86		2.65	
Annual MWh	3,779,310		4,709,015		2,002,088		2,858,153	
Cost Component	**M$/yr**	**$/ton CO_2***	**M$/yr**	**$/ton CO_2***	**M$/yr**	**$/ton CO_2***	**M$/yr**	**$/ton CO_2***
Annual Fixed Cost	$21.06	$5.97	$10.91	$2.49	$10.70	$5.75	$18.28	$6.89
Annual Variable Cost	$148.67	$42.17	$160.28	$36.61	$79.55	$42.78	$119.78	$45.12
Total Annual O&M Cost	$169.74	$48.15	$171.20	$39.11	$90.25	$48.54	$138.05	$52.01
Annualized Capital Cost	$70.10	$19.88	$58.30	$13.32	$36.02	$19.37	$55.64	$20.96
Total Levelized Annual Cost	$239.84	$68.03	$229.60	$52.45	$126.29	$67.92	$193.69	$72.97
Plant Name	**Heskett**		**Milton R. Young**		**Stanton**		**Total**	
Gross Electrical Output (MWg)	110		675		200		4051	
Amine Scrubber Use (MW)	26.8		164		48.6		984	
Wet Scrubber Add./Unit	Yes/Units 1 and 2		Yes/Unit 1		Yes/Unit 1		N/A	
Wet FGD Use (MW)	3.66		7.559		4.655		80.9	
Net Electrical Output (MW)	70.8		448		133		2682	
CO_2 Captured, Mtons/yr	0.45		2.73		0.81		16.4	
Annual MWh	465,444		2,944,285		874,672		17,632,967	
Cost Component	**M$/yr**	**$/ton CO_2***	**M$/yr**	**$/ton CO_2***	**M$/yr**	**$/ton CO_2***	**M$/yr**	**$/ton CO_2****
Annual Fixed Cost	$10.90	$24.42	$12.64	$4.63	$8.66	$10.70	$93.16	$8.69
Annual Variable Cost	$34.86	$78.09	$115.95	$42.49	$56.30	$69.59	$715.38	$50.98
Total Annual O&M Cost	$45.76	$102.50	$128.60	$47.13	$64.96	$80.30	$808.56	$59.67
Annualized Capital Cost	$18.78	$42.07	$45.95	$16.84	$22.56	$27.88	$307.35	$22.90
Total Levelized Annual Cost	$64.55	$144.59	$174.55	$63.96	$87.52	$108.18	**$1,116**	**$82.59**

* Captured.
** Average.

Table 20. Performance Comparison of the Two Technologies Analyzed

Units	PERFORMANCE SUMMARY	Subcritical PC Plant Without CO_2 Removal	Amine Process CO_2 Removal	Oxygen Combustion CO_2 Removal
% HHV	Net Therm. Efficiency	32.8%	20.1%	25.2%
Btu/kWh	Heat Rate (HHV)	10,393	16,954	13,521
MW	Gross Power	500	500	500
MW	Internal Power	41	218	166
lb/h	Fuel required	765,224	765,224	723,157
	EFFLUENTS			
lb/MWh	SO_2	0.33	0.00	24.27
lb/MWh	NO_x (NO_2)	0.29	0.22	0.00
lb/MWh	CO_2	3,543	354	198
lb/MWh	Particulates	0.23	0.15	0.05
lb/MWh	Solid Waste	305	161	178
%	SO_2 Removal	99%	99.97%	68%
%	NO_x Removal	80%	85%	100%
%	CO_2 Removal	0%	90%	93%
TPY	SO_2	229	1	26,625
TPY	NO_x (NO_2)	272	204	0
TPY	CO_2	3,280,245	328,024	216,994
TPY	Particulates	139	139	53
TPY	Solid Waste	142,956	148,628	195,771

Table 21. Cost Comparison for the Technologies Analyzed

Levelized Annual Costs (2005 US$)	Amine Scrubbing	Oxy-Fuel
Levelized Fixed Charges	$40,431,759	$77,843,171
Levelized Operating Costs	$74,683,199	$58,071,378
Total Annual Cost, $	$115,114,958	$135,914,549
Total Annual Cost, $/kW	$230.23	$271.83
$/ton CO_2 Captured	$37.20	$44.78
Increased Cost of Electricity COE, $/MWh	$62.17	$63.12
Cost of CO_2 Avoided, $/ton	$64.30	$63.87
Total Cost of Electricity, $/kWh	$0.108	$0.109

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