MEMBRANE PROCESS TO CAPTURE CO_2
FROM COAL-FIRED POWER PLANT FLUE GAS

FINAL REPORT

Award Number DE-NT0005312

Report Period: October 1, 2008 – March 31, 2011

by

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ABSTRACT

This final report describes work conducted for the U.S. Department of Energy National Energy Technology Laboratory (DOE NETL) on development of an efficient membrane process to capture carbon dioxide (CO_2) from power plant flue gas (award number DE-NT0005312). The primary goal of this research program was to demonstrate, in a field test, the ability of a membrane process to capture up to 90% of CO_2 in coal-fired flue gas, and to evaluate the potential of a full-scale version of the process to perform this separation with less than a 35% increase in the levelized cost of electricity (LCOE). Membrane Technology and Research (MTR) conducted this project in collaboration with Arizona Public Services (APS), who hosted a membrane field test at their Cholla coal-fired power plant, and the Electric Power Research Institute (EPRI) and WorleyParsons (WP), who performed a comparative cost analysis of the proposed membrane CO_2 capture process.

The work conducted for this project included membrane and module development, slipstream testing of commercial-sized modules with natural gas and coal-fired flue gas, process design optimization, and a detailed systems and cost analysis of a membrane retrofit to a commercial power plant.

The Polaris™ membrane developed over a number of years by MTR represents a step-change improvement in CO_2 permeance compared to previous commercial CO_2-selective membranes. During this project, membrane optimization work resulted in a further doubling of the CO_2 permeance of Polaris membrane while maintaining the CO_2/N_2 selectivity. This is an important accomplishment because increased CO_2 permeance directly impacts the membrane skid cost and footprint: a doubling of CO_2 permeance halves the skid cost and footprint. In addition to providing high CO_2 permeance, flue gas CO_2 capture membranes must be stable in the presence of contaminants including SO_2. Laboratory tests showed no degradation in Polaris membrane performance during two months of continuous operation in a simulated flue gas environment containing up to 1,000 ppm SO_2.

A successful slipstream field test at the APS Cholla power plant was conducted with commercial-size Polaris modules during this project. This field test is the first demonstration of stable performance by commercial-sized membrane modules treating actual coal-fired power plant flue gas.

Process design studies show that selective recycle of CO_2 using a countercurrent membrane module with air as a sweep stream can double the concentration of CO_2 in coal flue gas with little energy input. This pre-concentration of CO_2 by the sweep membrane reduces the minimum energy of CO_2 separation in the capture unit by up to 40% for coal flue gas. Variations of this design may be even more promising for CO_2 capture from NGCC flue gas, in which the CO_2 concentration can be increased from 4% to 20% by selective sweep recycle.

EPRI and WP conducted a systems and cost analysis of a base case MTR membrane CO_2 capture system retrofitted to the AEP Conesville Unit 5 boiler. Some of the key findings from this study and a sensitivity analysis performed by MTR include:

- The MTR membrane process can capture 90% of the CO_2 in coal flue gas and produce high-purity CO_2 (>99%) ready for sequestration.
- CO_2 recycle to the boiler appears feasible with minimal impact on boiler performance; however, further study by a boiler OEM is recommended.
- For a membrane process built today using a combination of slight feed compression, permeate vacuum, and current compression equipment costs, the membrane capture process can be competitive with the base case MEA process at 90% CO_2 capture from a coal-fired power plant. The incremental LCOE for the base case membrane process is about equal to that of a base case MEA process, within the uncertainty in the analysis.
- With advanced membranes (5,000 gpu for CO_2 and 50 for CO_2/N_2), operating with no feed compression and low-cost CO_2 compression equipment, an incremental LCOE of $33/MWh at 90% capture can be achieved (40% lower than the advanced MEA case).
- Even with lower cost compression, it appears unlikely that a membrane process using high feed compression (>5 bar) can be competitive with amine absorption, due to the capital cost and energy consumption of this equipment. Similarly, low vacuum pressure (<0.2 bar) cannot be used due to poor efficiency and high cost of this equipment.
- High membrane permeance is important to reduce the capital cost and footprint of the membrane unit. CO_2/N_2 selectivity is less important because it is too costly to generate a pressure ratio where high selectivity can be useful.
- A potential cost "sweet spot" exists for use of membrane-based technology, if 50-70% CO_2 capture is acceptable. There is a minimum in the cost of CO_2 avoided/ton that membranes can deliver at 60% CO_2 capture, which is 20% lower than the cost at 90% capture. Membranes operating with no feed compression are best suited for lower capture rates.

Currently, it appears that the biggest hurdle to use of membranes for post-combustion CO_2 capture is compression equipment cost. An alternative approach is to use sweep membranes in parallel with another CO_2 capture technology that does not require feed compression or vacuum equipment. Hybrid designs that utilize sweep membranes for selective CO_2 recycle show potential to significantly reduce the minimum energy of CO_2 separation.

EXECUTIVE SUMMARY

This final report describes work conducted for the U.S. Department of Energy National Energy Technology Laboratory (DOE NETL) on development of an efficient membrane process to capture carbon dioxide (CO_2) from power plant flue gas (award number DE-NT0005312). The work was conducted by the project partners from October 1, 2008 through March 31, 2011.

The primary goal of this research program was to demonstrate, in a field test, the ability of a membrane process to capture up to 90% of CO_2 in coal-fired flue gas, and to evaluate the potential of a full-scale version of the process to perform this separation with less than a 35% increase in the levelized cost of electricity (LCOE). Membrane Technology and Research (MTR) conducted this project in collaboration with Arizona Public Services (APS), who hosted a membrane field test at their Cholla coal-fired power plant, and the Electric Power Research Institute (EPRI) and WorleyParsons (WP), who performed a comparative cost analysis of the proposed membrane CO_2 capture process.

CO_2 capture from power plant flue gas is difficult for all separation technologies, including membranes, because of the low partial pressure of CO_2 in flue gas. In previous DOE NETL-funded work (DE-FC26-07NT43085), MTR made two innovations to address the challenges of CO_2 capture from power plant flue gas with membranes:

- New membranes with CO_2 permeances approximately tenfold higher than commercial CO_2-selective membranes were developed (CO_2 permeance = 1,000 gpu for new membranes compared with 100 gpu for conventional membranes used in natural gas processing). The high permeance of these new membranes – designated Polaris™ – will reduce the required membrane area, footprint, and capital cost of a membrane CO_2 capture system by nearly an order of magnitude.
- A novel process design that uses incoming combustion air as a sweep in a countercurrent module to generate separation driving force was proposed. This selective CO_2 recycle design reduces the need for costly and energy-intensive compression equipment.

These developments showed promise to yield an economically attractive membrane-based flue gas CO_2 capture process. The follow-on work described in this report included membrane and module development, slipstream testing of commercial-sized modules with coal-fired flue gas, process design optimization, and a detailed systems and cost analysis of a membrane retrofit to a commercial power plant. Key results from this program are summarized below.

The Polaris membrane represents a step-change improvement in CO_2 permeance compared to previous commercial CO_2-selective membranes. During this project, membrane optimization work resulted in a further doubling of the CO_2 permeance of Polaris membrane while maintaining the CO_2/N_2 selectivity. This is an important accomplishment because increased CO_2 permeance directly impacts the membrane skid cost and footprint: a doubling of CO_2 permeance halves the skid cost and footprint. In addition to providing high CO_2 permeance, flue gas CO_2 capture membranes must be stable in the presence of contaminants including SO_2. Laboratory tests showed no degradation in Polaris membrane performance during two months of continuous operation in a simulated flue gas environment containing up to 1,000 ppm SO_2.

In early 2010, a field test system designed to capture 1 ton CO_2/day (1 TPD) was installed at the APS Cholla coal-fired power plant. This system treated a slipstream of post-FGD flue gas and ran for three months during summer 2010. The test unit demonstrated stable Polaris module performance, as well as successful countercurrent operation with air as a sweep stream. To our knowledge, this field test was the first demonstration of commercial-sized membrane modules treating actual coal-fired power plant flue gas.

Process design studies show that selective recycle of CO_2 using a countercurrent membrane module with air as a sweep stream lowers the minimum energy required for CO_2 capture. A simple version of this design in which a generic capture unit is placed in series with a sweep membrane is shown in Figure ES1. The sweep membrane unit can double the concentration of CO_2 in coal flue gas with little energy input (the only energy use is for fans to push gas through the sweep membrane unit). This pre-concentration of CO_2 by the sweep membrane reduces the minimum energy of CO_2 separation in the capture unit by up to 40% for coal flue gas. Variations of this design may be even more promising for CO_2 capture from NGCC flue gas, in which the CO_2 concentration can be increased from 4% to 20% by selective sweep recycle.

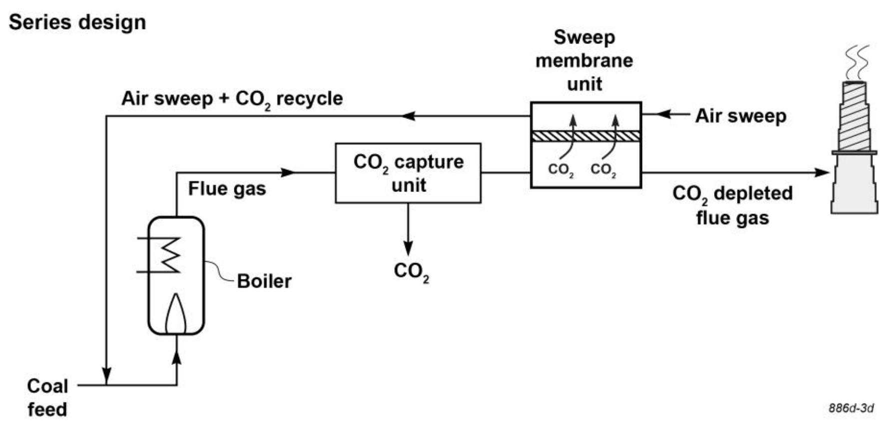

Figure ES1. Process schematic of a selective recycle sweep membrane unit operating in series with a generic CO_2 capture unit for treating coal flue gas. The capture unit can be membrane-based, or can use another separation technology such as absorption. For the cost analysis performed in this report, the capture unit uses a membrane with a permeate vacuum.

EPRI and WP conducted a systems and cost analysis of a base case MTR membrane CO_2 capture system retrofitted to the AEP Conesville Unit 5 boiler. The Figure ES1 process design was used, in which the first step capture unit was a cross-flow membrane operating with slight feed compression (2.0 bar) and a vacuum on the permeate (0.2 bar). The base case analysis used current membrane properties. Some of the key findings from this study and a sensitivity analysis performed by MTR include:

- The MTR membrane process can capture 90% of the CO_2 in coal-derived flue gas and produce high-purity CO_2 (>99%) ready for sequestration.

- CO_2 recycle to the boiler appears feasible with minimal impact on boiler performance, based on a preliminary evaluation by EPRI/WP; however, because this selective recycle is key to the viability of the membrane process, further study by a boiler OEM is recommended, preferably through testing on a small-scale boiler.

- For a membrane system built today using a combination of slight feed compression (2.0 bar), permeate vacuum (0.2 bar), and current compression equipment costs ($2,150/kW), the membrane capture process can be competitive with the base case MEA process at 90% CO_2 capture from a coal-fired power plant. The EPRI/WP analysis indicates the membrane process will have capital cost about 25% more than that of the MEA process (with 30% uncertainty), but will use less energy and have lower O&M costs. The overall effect is that the incremental LCOE for the base case membrane process is about equal to that of a base case MEA process ($56/MWh), within the uncertainty in the analysis. A comparison of incremental LCOE for the amine and the membrane processes is shown in Figure ES2.

- For the base case membrane system, with compression equipment priced at $2,150/kW, compression costs make up more than 50% of the membrane CO_2 capture system costs. Lower compression costs, such as those used in the recent DOE Bituminous Baseline report ($1,030/kW), will dramatically improve the competitiveness of a membrane process. For example, low-cost compression equipment will lower the membrane incremental LCOE to $43/MWh (see Figure ES2). It seems likely that in the time horizon predicted for commercialization of CO_2 capture technology (deployable by 2020), lower compression costs are possible.

- Even with low-cost compression, it appears unlikely that a membrane process using high feed compression (>5 bar) can be competitive with amine absorption, due to the capital cost and energy consumption of this equipment. Similarly, low vacuum pressure (<0.2 bar) cannot be used due to poor efficiency and high cost of this equipment. For these reasons, the MTR membrane design uses a feed pressure of 1 – 2 bar and a permeate pressure of ≥0.2 bar.

- High membrane permeance is important to reduce the capital cost and footprint of the membrane unit. CO_2/N_2 selectivity is less important because it is too costly to generate a pressure ratio where high selectivity can be useful. We recommend that membrane targets be set at 5,000 gpu for CO_2 and 50 for CO_2/N_2. Advanced membranes with these properties, operating with no feed compression and low-cost CO_2 compression equipment, would produce an incremental LCOE of $33/MWh at 90% capture (40% lower than the advanced MEA case).

- A potential cost "sweet spot" exists for use of membrane-based technology, if 50-70% CO_2 capture is acceptable. There is a minimum in the cost of CO_2 avoided/ton that membranes can deliver at 60% CO_2 capture, which is 20% lower than the cost at 90% capture. Membranes operating with no feed compression are best suited for lower capture rates, because without compression, the remaining CO_2 driving force is very low at high capture rates. As a result, the membrane area (and cost) increase rapidly at capture rates >70%.

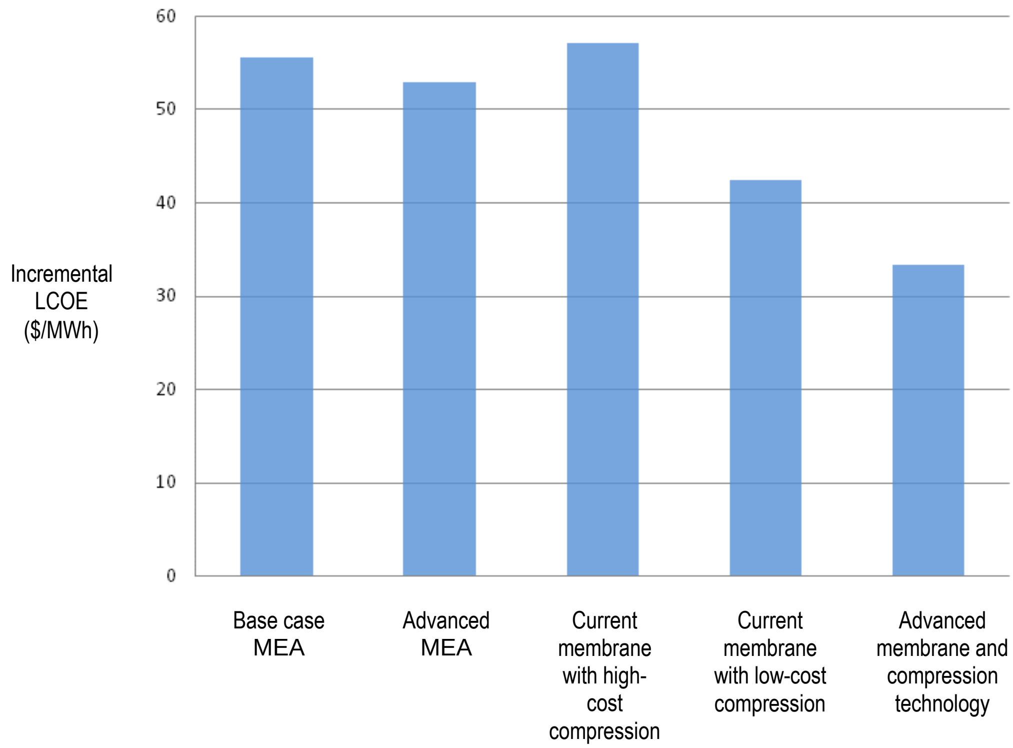

Figure ES2. Incremental LCOE for 90% CO_2 capture from coal power plants retrofitted with various capture technologies.

Currently, it appears that the biggest hurdle to use of membranes for post-combustion CO_2 capture is compression equipment cost. An all-membrane capture process must use some flue gas feed compression and/or a permeate vacuum to generate a driving force for CO_2 separation. Because CO_2 compression is a focus of CCS research, it seems likely that more-efficient, more-affordable compression equipment will be available when CO_2 capture from power plants is eventually commercialized. However, this may not be the case for large vacuum equipment because it is not required by other capture technologies (such as amine absorption). Future work will need to clarify the potential for vacuum cost and performance improvements.

An alternative approach that would reduce the reliance on compression equipment developments is to use sweep membranes in parallel with another CO_2 capture technology that does not require feed compression or vacuum equipment. Hybrid designs that utilize sweep membranes for selective CO_2 recycle show potential to significantly reduce the minimum energy of CO_2 separation. This is particularly true for NGCC flue gas where sweep membranes can pre-concentrate CO_2 from 4% to 20% with little energy input. In addition to continued development of all-membrane capture processes, we believe such hybrid designs warrant further examination.

Finally, in addition to the base case systems and cost analysis conclusions, membrane systems have a number of unique attributes that may be beneficial for use in post-combustion CO_2 capture systems. Examples include:

- Membranes are not poisoned by SO_2, and in fact will remove SO_2 from flue gas even more efficiently than they will remove CO_2. This suggests the possibility of co-capture of SO_2 and CO_2 by membranes. If co-capture is viable, cost savings associated with replacing two unit operations (FGD for SO_2 removal and a CO_2 capture technology) with a single membrane system could be realized.
- As a passive separation process, membranes do not emit VOCs or require hazardous chemicals handling or disposal. This may make permitting issues related to installation of the membrane capture system much easier compared to amine absorption. Amine systems require daily handling and disposal of large amounts of make-up amine solution, and must address additional concerns about emissions of nitrosamines and nitramides as by-products in the flue gas.
- The modular nature of membrane separation units are a potential advantage if different CO_2 capture rates are required over the life of the plant (for example, a 50% capture rate that is later increased to 90% to adjust to progressive regulations). In this scenario, additional membrane modules can simply be added when needed.

These potential benefits and their impact on the competitiveness of a membrane CO_2 capture system have not been quantified in this report. We recommend that future comparative assessments examine these items in more detail to clarify their importance in a competitive technology environment.

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1. INTRODUCTION

Background

During the last century, the concentration of carbon dioxide (CO_2) in the atmosphere increased from 275 to 387 ppm. This increase is largely due to the combustion of fossil fuels and has already produced measurable increases in global temperatures. Figure 1 illustrates the correlation between man-made CO_2 emissions resulting from fossil fuel combustion and atmospheric CO_2 concentration over the past 250 years. Climate models indicate that continuation of this trend will dramatically change the global climate by 2100.[1]

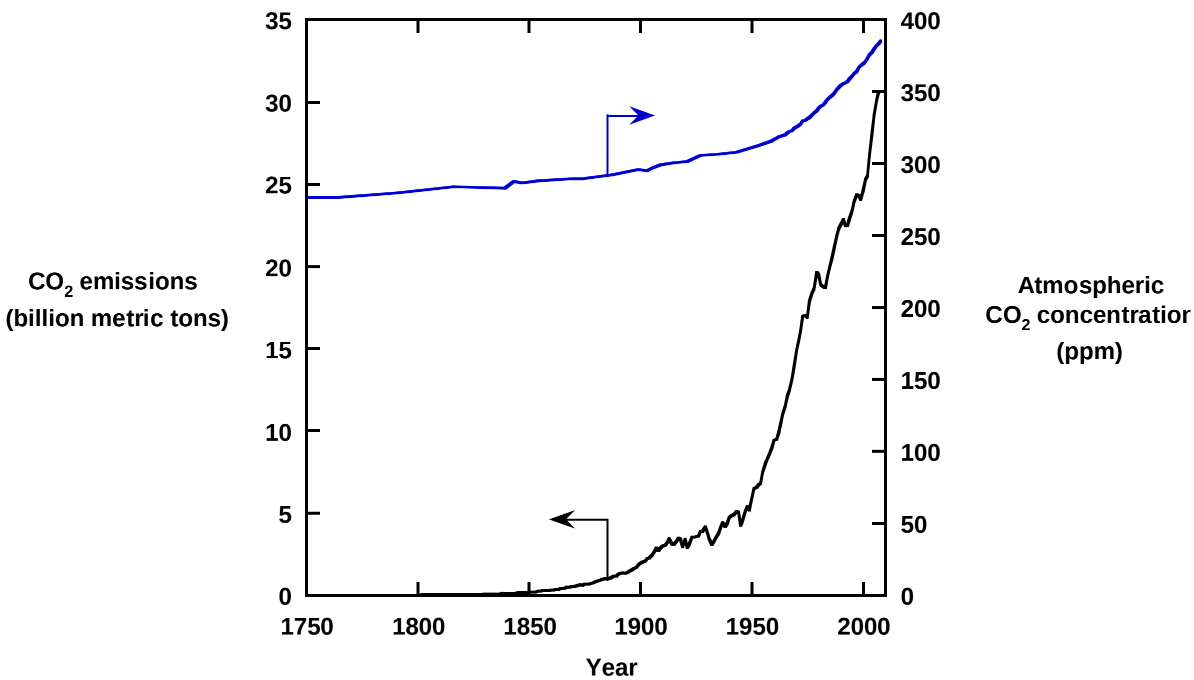


Figure 1. Anthropogenic CO_2 emissions and atmospheric CO_2 concentration over the past 250 years. Data are from the Oak Ridge National Laboratory, Carbon Dioxide Information Center.

Coal-fired power plants generate more than 50% of the electricity in the United States and produce about 40% of the country's CO_2 emissions. Globally, the situation is similar. Because of the relatively low cost and large domestic supply of coal, power production from this fuel is expected to increase over the next 20 years.[2,3] According to the Energy Information Agency, growing power demands will result in a 50% increase in installed coal-fired electricity generating capacity by 2030.[4] One way to reduce CO_2 emissions to the atmosphere is carbon capture and sequestration (CCS). In this scheme, CO_2 is captured from large point sources, such as power plants, and sequestered underground in geological formations for long periods of time. A key to this approach is technology that can separate CO_2 from process gases cost effectively, allowing it to be sequestered without radically increasing energy costs.

Three pathways for CO_2 capture from fossil fuel power production are being considered by researchers: post-combustion CO_2 capture from flue gas, pre-combustion capture from syngas, and oxy-combustion, which produces a nearly sequestration-ready CO_2 effluent.[3] These approaches to decarbonized energy production are shown schematically in Figure 2. Oxy-combustion and gasification processes, such as the Integrated Gasification Combined Cycle (IGCC) process, are new methods of producing energy from fossil fuels that generate by-product CO_2 in a relatively concentrated or high partial pressure stream. This makes CO_2 capture from these processes relatively easy. However, there is very little commercial experience with IGCC or oxy-combustion power production, and essentially all of the existing fossil fuel electricity in the U.S. is produced via conventional combustion in air. The majority of these existing plants produce electricity from combustion of pulverized coal. If power plant CO_2 emissions are to be addressed, post-combustion CO_2 capture technology will have to be applied to these existing coal plants.

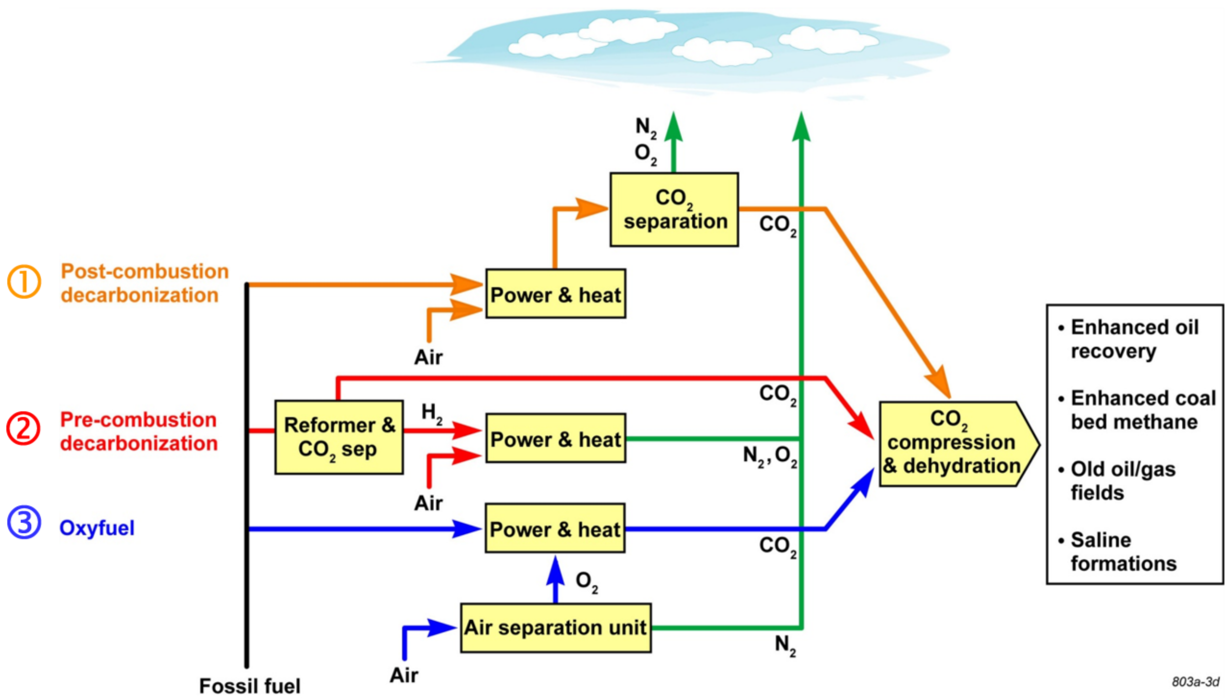


Figure based on information from: J.D. Figueroa *et al., International Journal of Greenhouse Gas Control 2*, 9-20 (2008)[3]

Figure 2. Routes to decarbonized fossil fuel power production.

Post-combustion capture of CO_2 from power plant flue gas has been the subject of many studies. Useful reviews include a recent IEA Greenhouse Gas Program Report,[5] MIT's *The Future of Coal*,[2] and a DOE NETL overview paper.[3] Currently, amine absorption is the leading candidate technology for use in post-combustion CO_2 capture, primarily because it is a proven technology used successfully to remove CO_2 from industrial gas streams for decades. However, a number of studies have shown that amine absorption, when applied to flue gas CO_2 capture, is going to be costly and energy intensive.[3,6] Alternative CO_2 capture technologies, such as absorption by chilled ammonia, are also being evaluated for flue gas treatment. Some processes show promise, but at this time there is no clear winning technology.

Membranes – a relatively new industrial gas separation technology – have also been suggested as a way to capture CO_2 from flue gas.[7-9] Membrane processes offer a number of advantages when applied to flue gas CO_2 capture, including low energy use, tolerance to flue gas contaminants (SO_x, NO_x, etc.), no use of harmful chemicals with disposal issues, recovery of flue gas water, and – because they use only electric power – no modifications to existing boilers and steam turbines. However, the Achilles heel of membrane processes has been the enormous membrane area required for separation, because of the low partial pressure of CO_2 in flue gas. MTR has made two key innovations to address this problem:

1. New membranes with ten times the CO_2 permeance of conventional gas separation membranes have been developed. A tenfold increase in permeance leads to a tenfold decrease in the required membrane area, and reduces the capital cost and footprint of the capture system substantially.
2. A selective flue gas recycle process has been developed. This process uses an existing air stream to generate a driving force for transmembrane CO_2 transport in a countercurrent/sweep module, reducing the need for compressors or vacuum pumps and the associated energy costs.

These innovations offer the potential for a membrane process to capture CO_2 from flue gas in a cost-effective manner. In prior work (DE-FC26-07NT43085), we have demonstrated bench-scale performance of high permeance CO_2 membranes with flue gas mixtures and the effective operation of countercurrent/sweep modules. Field work for this project tested 8-inch-diameter membrane modules with small slipstreams of flue gas from coal-fired power plants. This report describes results from bench and slipstream membrane testing, as well as a systems and economic analysis of the MTR CO_2 capture membrane process applied to a commercial-scale coal-fired power plant.

Membrane Fundamentals

Polymer membranes separate the components of a gas or vapor mixture because the components permeate the membrane at different rates. Gas flux, N_A (cm^3(STP)/s), through a nonporous polymeric membrane can be expressed as follows:[10]

$$N_A = \frac{P_A}{l} \cdot \Delta p_A \cdot A \tag{1}$$

where P_A (cm^3(STP) cm/cm^2·s·cmHg) is the permeability coefficient of gas component A, l is the thickness of the membrane selective layer [cm], Δp_A is the partial pressure difference across the membrane (cmHg), and A is the required membrane area (cm^2). The permeance, P_A/l (cm^3(STP)/cm^2·s·cmHg), is also used to characterize the pressure-normalized flux in composite membranes where the thickness of the selective layer cannot be measured.

If the diffusion process obeys Fick's law and the downstream pressure is much less than the upstream pressure, permeability is given by:

$$P_A = D_A \times S_A \tag{1}$$

where D_A is the average effective diffusivity (a measure of the gas mobility), and S_A is the solubility of penetrant A in the polymer, which links the concentration of the gas in the membrane to the pressure in the adjacent gas. The separating ability of a membrane is determined by the selectivity, α, defined as the ratio of the gas permeabilities, P_A/P_B, or permeances. Selectivity can be expressed as

$$\alpha_{A/B} = \frac{P_A}{P_B} = \left(\frac{D_A}{D_B}\right) \times \left(\frac{S_A}{S_B}\right) \tag{1}$$

where D_A/D_B is the diffusivity selectivity, which is the ratio of the diffusion coefficients of gases A and B. The ratio of the solubility of gases A and B, S_A/S_B, is the solubility selectivity. In glassy polymers, the dominant contribution to selectivity is the diffusivity selectivity, which depends on the ratio of the molecular sizes. In rubbery polymers, the dominant contribution is from the solubility selectivity, which is proportional to the ratio of the permeant condensabilities. CO_2 is both smaller than nitrogen and much more condensable, so membranes are always selective for CO_2 over N_2 to varying degrees. All membranes used commercially for industrial gas separations, including the CO_2 capture membranes developed by MTR, operate by the solution-diffusion mechanism described above.

Membrane Development at MTR

Conventional membranes cannot capture CO_2 from flue gas economically because the low partial pressure of CO_2 in flue gas, combined with the enormous gas flow rates of power plants, require prohibitively large membrane areas. Our design calculations show that membranes with CO_2 permeance of at least 1,000 gpu (1 gpu=10^{-6} cm^3 (STP)/$cm^2 \cdot s \cdot cmHg$) are needed to make CO_2 capture with membranes attractive, a value ten times higher than that of current commercial CO_2 separation membranes.

Recently, with DOE NETL support, MTR has developed new membranes specifically designed for flue gas CO_2 capture. These membranes – named Polaris™ – are based on hydrophilic polymers and are extremely permeable to CO_2 and polar species such as water, NO_x and SO_x. Because these membranes transport molecules by simple passive solution-diffusion, they are inert to flue gas components such as water, oxygen, SO_x, and NO_x. Figure 3 shows a trade-off plot of CO_2/N_2 selectivity versus CO_2 permeance for MTR Polaris membranes. Compared to a typical commercial CO_2-selective membrane, Polaris membranes are substantially more permeable and have better selectivity. For example, the Polaris membranes have CO_2/N_2 selectivity ranging from 50 to 60 and a CO_2 permeance of ~1,000 gpu. For comparison, a good cellulose acetate membrane used for removing CO_2 from natural gas has a CO_2 permeance of around 100 gpu combined with a CO_2/N_2 selectivity of 30.

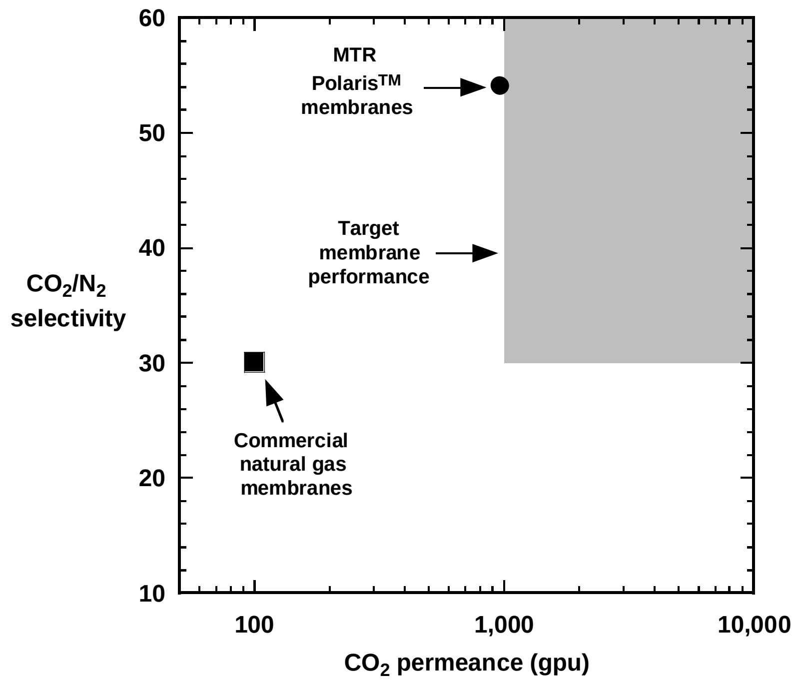


Figure 3. A CO_2/N_2 trade-off plot showing data for MTR Polaris™ compared with the properties of a good commercial natural gas membrane. The shaded region is the membrane performance target area necessary for an economic CO_2 capture process. Data are pure-gas values at room temperature.

As shown in Figure 3, the transport properties of the Polaris membranes extend into the target window identified from process simulations as the performance levels necessary for an economic CO_2 capture process. Several of the Polaris membrane formulations have been scaled up for production on MTR's commercial casting and coating equipment, and fabricated into commercial-sized modules. Additional improvements in membrane performance are possible, and important, to further reduce the capital cost of a membrane CO_2 capture process. Details of membrane development are discussed in Section 2 of this report.

Membrane Process Design

A major challenge for post-combustion CO_2 capture membrane systems is generation of the partial pressure driving force required to separate CO_2 from the flue gas. Typically, a post-combustion membrane system is placed downstream of the flue gas desulfurization (FGD) unit in the same location as that being proposed for amine scrubbing. At this point, the flue gas stream is just slightly above atmospheric pressure and contains perhaps 10-13% CO_2 for a coal-fired power plant, with the bulk being mostly nitrogen.

Figure 4 illustrates the two simplest membrane process designs for CO_2 capture from power plant flue gas. In these single-stage membrane processes, flue gas is fed to a membrane module, and a pressure driving force is generated by either (a) compression on the feed side or (b) a vacuum on the permeate side of the membrane. For both cases, 90% of the flue gas CO_2 is captured in the membrane permeate stream, and the pressure ratio across the membrane is

5.5. This pressure ratio was chosen because it corresponds to the value obtained when using no feed compression (1.1 bar) and the minimum practical vacuum pressure (0.2 bar). Calculations show that the required energy is lower for the vacuum process because the vacuum only has to pump about 10% of the flue gas that permeates the membrane (largely CO_2), whereas a feed compressor pressurizes all of the flue gas (CO_2 plus the bulk N_2). Perhaps more importantly, a vacuum process is likely to be cheaper than feed compression because pressurizing the flue gas feed requires not only large compression equipment but also enormous turboexpanders to recover energy. For example, although the net power use for the feed compression case shown in Figure 4(a) is 69 MW_e, this process actually uses 123 MW_e of compression and recovers 52 MW_e with turboexpanders. The capital cost of the feed compression case will scale with the gross power (175 MW_e), and will be significantly more costly than for the vacuum case (56 MW_e).

On the other hand, a vacuum process will require a larger membrane area than compression, because the CO_2 partial pressure difference across the membrane is small. The overall balance between these factors – discussed in detail in Section 6 of this report – favors minimal feed compression as the lowest-cost approach to post-combustion CO_2 capture with membranes. This conclusion dictates that very permeable membranes are required to minimize the membrane system footprint and cost.

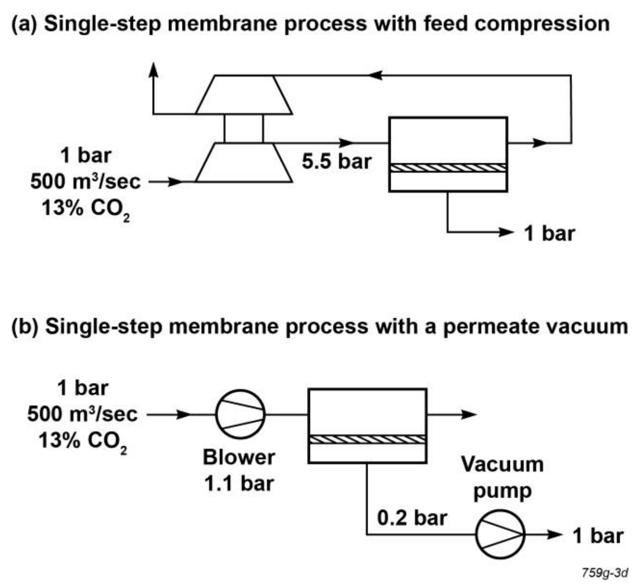


Figure 4. Single-step membrane processes to capture CO_2 in flue gas using (a) feed compression and (b) permeate vacuum at a 600 MW_e power plant. Process simulations show that for (a), the membrane area is 1.9 million m^2 and the net power use is 69 MW_e; for (b), the membrane area is 9.7 million m^2 and the power use is 56 MW_e. Both processes capture 90% of the CO_2 in the flue gas using membranes with CO_2 permeance of 1,000 gpu and CO_2/N_2 selectivity of 50.

In addition to large membrane area and power requirements, single-stage membrane designs are unable to produce high-purity CO_2 *combined* with high CO_2 capture (here and throughout, CO_2 capture refers to the amount of CO_2 that permeates the membrane divided by the amount in the feed to the membrane). In fact, a single-stage membrane process alone cannot produce high-purity CO_2 in the permeate with 90% CO_2 capture, regardless of the membrane selectivity. This is because the system performance is limited by the pressure ratio across the membrane.

The importance of pressure ratio in the separation of gas mixtures can be illustrated by considering the separation of a gas mixture with component concentrations (mole fractions) y_{i_o} and y_{j_o} at a feed pressure of p_o. A flow of component *i* across the membrane can only occur if the partial pressure of component *i* on the feed side of the membrane, $p_{i_o} = y_{i_o} p_o$, is greater than the partial pressure of component *i* on the permeate side of the membrane, $p_{i_l} = y_{i_l} p_l$. That is, permeation only occurs if $p_{i_o} > p_{i_l}$ or $y_{i_o} p_o > y_{i_l} p_l$. It follows that the maximum separation achieved by the membrane can be expressed as

$$\frac{\text{Permeate concentration}}{\text{Feed concentration}} = \frac{y_{i_l}}{y_{i_o}} \leq \frac{p_o}{p_l} = \frac{\text{Feed pressure}}{\text{Permeate pressure}} \tag{1}$$

This means that the separation achieved can never exceed the pressure ratio of p_o / p_l, no matter how selective the membrane. In practical separation applications, the pressure ratio across the membrane is usually between 5 and 15. Higher pressure ratios can be achieved by using larger compressors on the feed gas or larger vacuum pumps on the permeate, but the capital and energy cost of this equipment limits the practical range. For flue gas treatment with a vacuum membrane system, positive displacement pumps can reach a theoretical suction pressure of 0.05 bar. However, accounting for leaks, the large permeate volumetric flow rate, pressure drops in tubing and module permeate channels, and the size and cost of vacuum equipment (which increases as the suction pressure decreases), we believe the lowest realistic pressure on the permeate side of the membrane will be about 0.2 bar. For a feed pressure slightly above atmospheric pressure (1.1 bar), this corresponds to a pressure ratio of only 5.5.

The membrane vacuum process shown in Figure 4(b) provides an example of the impact of pressure ratio on CO_2 capture from flue gas. In this case, the feed-to-permeate pressure ratio is 5.5 (1.1 bar/0.2 bar). Under these conditions, the difference in performance for a membrane with a selectivity of 50 or one with selectivity of 500 is small. This point is illustrated in Figure 5, which shows the permeate CO_2 concentration as a function of permeate pressure for membranes with these selectivities. The calculations were performed using a computer simulation program (ChemCad 5.6, ChemStations, Houston, TX) containing code for the membrane operation developed by the MTR Engineering Group. In these simulations, the CO_2 capture rate is fixed at 90%.

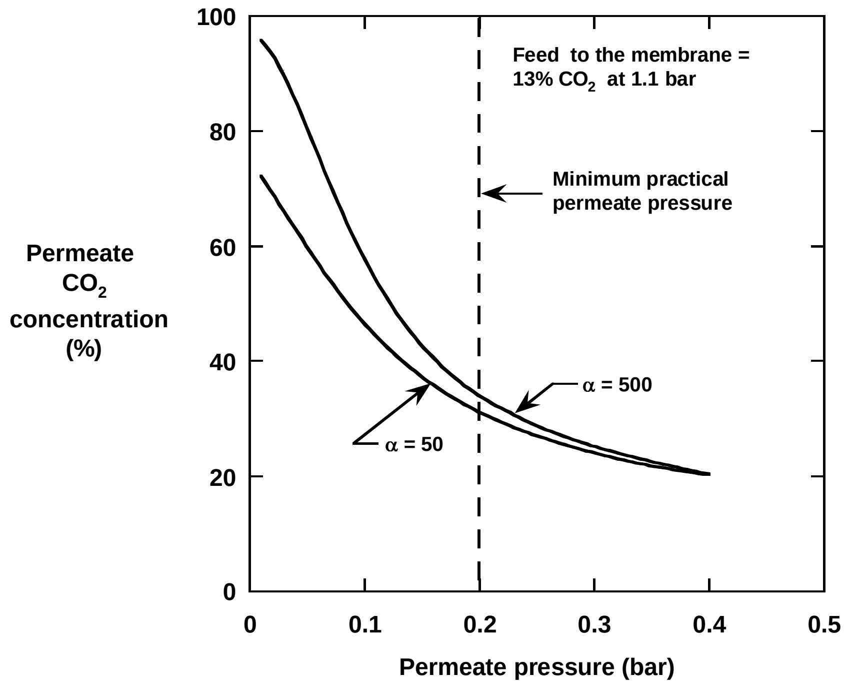
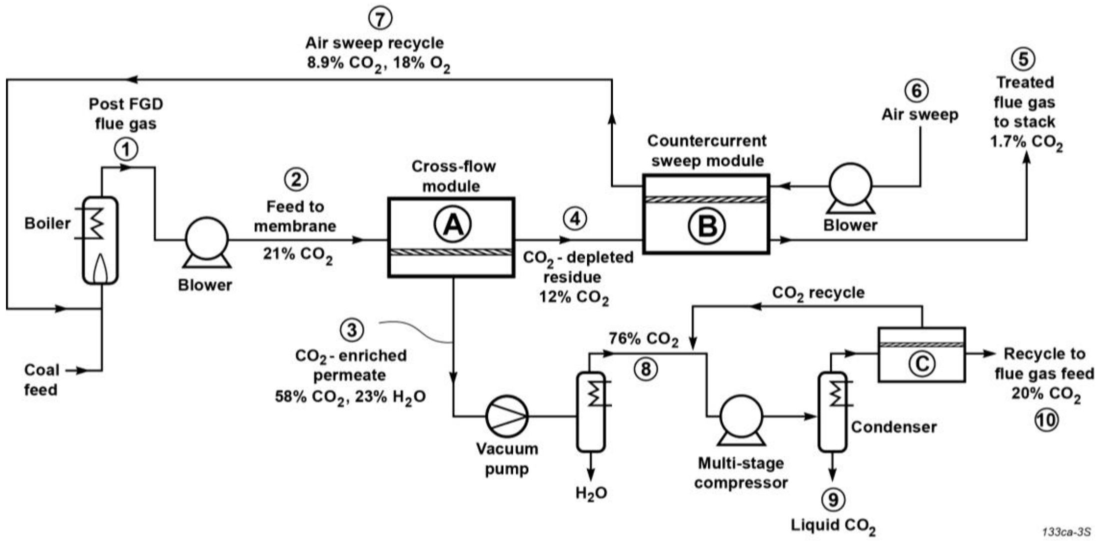


Figure 5. Calculated permeate CO_2 concentration (mol%) as a function of permeate pressure for membranes with CO_2/N_2 selectivities of 50 and 500. CO_2 capture was fixed at 90% so that the gas leaving the membrane in the residue contains about 2% CO_2.

At the lowest practical permeate pressure (0.2 bar), the difference in CO_2 permeate concentration produced by the two membranes is insignificant. Only at pressures below the minimum practical operating pressure is there a significant benefit to using the high selectivity membrane. In addition, even if it was affordable to generate low vacuum pressures, a large increase in selectivity is inevitably associated with a large decrease in membrane permeance. Reduced membrane permeance means larger membrane area is required. For example, at 90% CO_2 capture with a pressure ratio of 5.5, a membrane with a CO_2/N_2 selectivity of 500 requires about 10 times the membrane area of one with a selectivity of 50.

We make this point about pressure ratio because there is a widespread belief in the engineering community that higher selectivity membranes are required for a useful CO_2 separation. In fact, the point of diminishing returns in membrane processes is typically reached when the selectivity is about three to five times the maximum practical pressure ratio (in this case, corresponding to a CO_2/N_2 selectivity of 15 to 30). In a pressure-ratio-limited case like flue gas CO_2 capture, high membrane permeance is much more important than high selectivity. Higher selectivity only increases the required membrane area without producing an improvement in product purity (concentration).

Because of these pressure ratio constraints, treatment of flue gas requires a multi-step or multi-stage membrane design. MTR has screened a wide variety of multi-stage/step designs to identify the most efficient process. Our current best design uses a radically different approach to generate driving force – countercurrent sweep using combustion air. In this scheme, air going to the boiler flows across the permeate side of the membranes, improving the partial pressure driving force across the membrane without changing the flue gas pressure. A process design incorporating this countercurrent/sweep concept is shown in Figure 6.



Component	Raw Flue Gas ①	Feed to First Membrane Step ②	First Membrane Step Permeate ③	First Membrane Residue ④	Treated Flue Gas ⑤
Composition (vol%)					
Carbon dioxide	21	21	58	12	1.7
Nitrogen	70	71	18	84	94
Oxygen	2.1	2.1	1.0	2.4	4.7
Water	7.7	6.2	23	2.0	-
Pressure (bar)	1.0	2.0	0.2	1.9	1.0
Flow rate (MMscfd)	1,660	1,640	320	1,320	1,100
Component	*Air Sweep to Membrane* ⑥	*Sweep Recycle to Boiler* ⑦	*Dry Permeate* ⑧	*Liquid CO_2* ⑨	*Recycle Gas* ⑩
Composition (vol%)					
Carbon dioxide	-	8.9	76	99.6	20
Nitrogen	79	72	23	-	76
Oxygen	21	17	1.3	-	4.3
Water	-	1.0	-	-	-
Pressure (bar)	1.1	1.0	9	27	27
Flow rate (MMscfd)	1,300	1,520	245	170	74

Figure 6. Simplified flow diagram of the proposed membrane process to capture CO_2 in flue gas from a 550 MW_e coal-fired power plant.

Flue gas from the boiler is sent to a first membrane step (unit A) and a vacuum pump is used on the permeate side of this membrane. Because the volume of the permeate gas (stream ③) passing through the vacuum pump is only a fraction of the volume of the flue gas (stream ②), the power used by the vacuum pump is much smaller than the power that would be consumed by compressing the feed gas to achieve the same pressure ratio. This cross-flow membrane unit removes a portion of the CO_2 in the flue gas, leaving a residue stream (stream ④) that still contains about 12% CO_2. This gas passes on one side of a second membrane (unit B) that has a countercurrent/sweep configuration. A portion of the feed air to the boiler (stream ⑥) passes on the other side of this membrane as a sweep stream. Because of the difference in concentration of CO_2, some CO_2 passes through the membrane and is recycled with the feed air to the boiler (stream ⑦). The treated flue gas (stream ⑤) leaving the countercurrent membrane unit contains about 2% CO_2 and is vented. Overall, 90% CO_2 capture is achieved.

The permeate gas leaving the vacuum pump contains about 76% CO_2 on a dry basis. This gas (stream ⑧) is sent to a compression-condensation-membrane loop. This type of energy-efficient loop is used commercially in the MTR VaporSep® process to recover hydrocarbon liquids in the petrochemical industry. The liquefaction section uses about 6% of the electric power made by the plant to deliver high-pressure supercritical CO_2 to the pipeline for sequestration (stream ⑨). A small recycle (stream ⑩) is blended with stream ① leaving the boiler. Overall, at 90% CO_2 capture, the membrane process uses about 12% of generated power to separate CO_2 from flue gas, plus 6% for the liquefaction loop. Additionally, according to current NETL projections, about 4% of power plant energy will be required for CO_2 transportation, storage, and monitoring costs. In total, therefore, this process uses slightly more than 20% of the power plant energy to capture and sequester 90% of the CO_2 in coal-fired power plant flue gas. This projected energy use compares favorably with other CO_2 capture technologies.

Compared to previous membrane designs considered for flue gas CO_2 capture, the MTR two-step selective recycle process offers a number of advantages:

1. Countercurrent/sweep with combustion air greatly reduces the energy required for 90% CO_2 capture. Membranes are generally considered to be an energy efficient way to do bulk separations; however, at high capture rates, membrane separation often becomes inefficient because driving force is lost. Consequently, a large fraction of the process energy and membrane capital cost is needed to remove the last few percent of the species being separated. In the Figure 6 design, the first membrane step – which uses vacuum to provide driving force – only needs to capture ~50% of the CO_2 in the initial pass. The remaining 40% of the CO_2 is removed in the countercurrent/sweep membrane step where combustion air on the permeate side of the membrane provides driving force that is essentially free (that is, no compression or vacuum required).
2. Recycling CO_2 to the boiler with a sweep membrane increases the concentration of CO_2 in the flue gas (from 12% to 18-24%), which improves driving force for membrane separation. By increasing the driving force for CO_2 capture, the membrane area (and capital cost) is reduced compared to a conventional membrane design that does not use selective CO_2 recycle. The impact of this recycle on boiler performance will be explored further in follow-on work.
3. The CO_2 purification section of the Figure 6 design uses an efficient membrane-assisted refrigeration process to produce high-purity liquid CO_2 ready for sequestration. This hybrid approach allows the overall process to produce CO_2 purities that could not be achieved with membranes alone. At the same time, the membranes pre-concentrate CO_2 so that the refrigeration process can operate at higher temperatures (using less costly materials) and with a simplified design.

Report Objectives and Organization

The purpose of this report is to document work conducted by MTR and our subcontractors, Arizona Public Services (APS), the Electric Power Research Institute (EPRI), and WorleyParsons (WP), to better understand the potential of membrane technology to be used for flue gas CO_2 capture. This work involved laboratory membrane/module development at MTR, slipstream field testing of membrane modules at APS, and a systems/economic analysis by EPRI, WP and MTR. The remainder of this report is organized in the following manner:

- Section 2 describes membrane and module development and testing at MTR;
- Section 3 reviews the membrane field tests conducted at the APS Cholla coal-fired power plant;
- Section 4 provides a minimum energy analysis of the selective CO_2 recycle concept;
- Section 5 discusses the energy use and cost of the membrane CO_2 capture process shown in Figure 6 (page 19), as evaluated by EPRI/WP;
- Section 6 presents a sensitivity analysis that examines the key factors affecting the membrane CO_2 capture process cost and performance, and
- Section 7 summarizes our conclusions and recommendations for future work.

Finally, the full comparative economic analysis prepared by the Electric Power Research Institute and WorleyParsons (EPRI/WP) is provided in the EPRI/WP Appendix of this report; their analysis is also summarized in the Section 5 discussion.

2. MEMBRANE AND MODULE DEVELOPMENT

Polaris^{TM} Membrane Development

Power plant flue gas has a low partial pressure of CO_2 and enormous volumetric flow rates. Even using the cost-effective two-step MTR process design, initial calculations showed that membranes must have a minimum CO_2 permeance of about 1,000 gpu (where 1 gpu = 10^{-6} cm^3 (STP)/ $cm^2 \cdot s \cdot cmHg$) and CO_2/N_2 selectivity of greater than 30 to make CO_2 capture with membranes economically feasible. In our previous DOE NETL project (DE-NT43085), Polaris^{TM} membranes with CO_2 permeances of 1,000 gpu and CO_2/N_2 selectivities of 50 were developed. In this project, we continued our efforts to optimize the membrane configuration, in an attempt to further increase CO_2 permeances and maintain CO_2/N_2 selectivities. There is a compelling reason to strive for higher membrane CO_2 permeance: doubling the CO_2 permeance will roughly halve the required membrane area, and thus significantly reduce the capital cost and footprint of a membrane CO_2 capture system.

Figure 7 illustrates the structure of a typical multilayer composite membrane. The microporous support material, with negligible resistance to gas permeation, provides mechanical strength to the membrane. The microporous support is often coated with a highly permeable gutter layer, which improves the compatibility between the support and selective layer, as well as conducting the permeating gas to the support membrane pores. The gutter layer is then coated with a selective layer composed of polymers with desirable properties for CO_2/N_2 separation. The overall separation performance of composite membranes largely depends on the properties of the selective layer, including its permeability, thickness, and integrity. Therefore, most of our effort has been focused on optimizing the selective layer structure and material properties, to achieve membranes with higher CO_2 fluxes and CO_2/N_2 selectivities. However, for flue gas CO_2 capture, because the membranes are very permeable (ten times higher CO_2 permeances than conventional membranes), modifications to the membrane support layers were also necessary to prevent resistance in these layers from adversely affecting membrane performance.

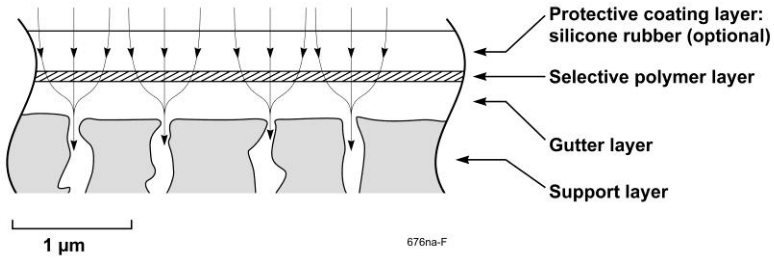


Figure 7. Schematic drawing of the structure of a composite membrane.

Figure 8 shows the progress of Polaris membrane development during this project ("2008 Baseline" and "Cholla I 2009"), as well as subsequent improvements achieved in a follow-on DOE-funded program ("End 2010" and "2011 Developmental membranes" results are from project DE-FE0005795). The data are presented in the form of a trade-off plot, where CO_2/N_2 selectivity is plotted against CO_2 permeance. Over time, we have steadily improved the performance of Polaris membranes, particularly by increasing CO_2 permeance (data points move to the right on the figure). Recently prepared Polaris membranes show CO_2 permeances of over 2,000 gpu, more than 100% higher than the original base case membrane, with similar CO_2/N_2 selectivities.

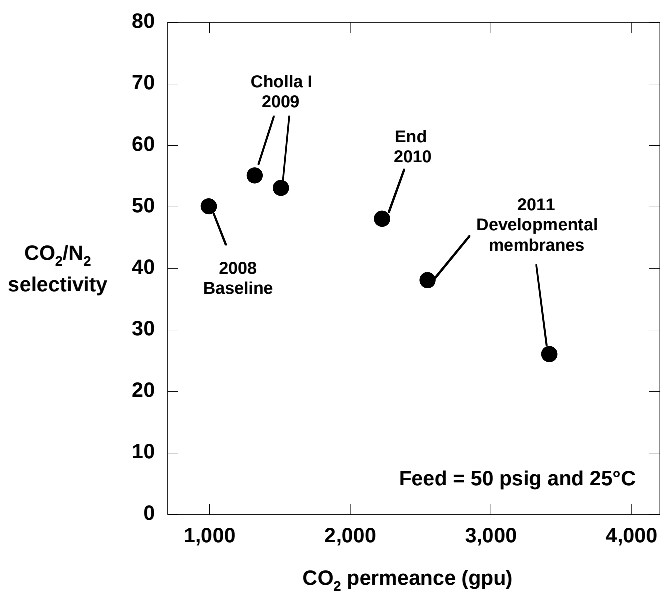


Figure 8. A CO_2/N_2 trade-off plot showing recent progress in Polaris membrane development.

Membranes with higher CO_2 permeances will always benefit the economics of a membrane-based CO_2 capture process. To get an idea of the upper limit for membrane performance, polymer materials with CO_2 permeabilities of more than 1,000 Barrer are known in the literature, and MTR can reliably make selective layer coatings of 0.1 µm thickness. Therefore, in principle, membranes with CO_2 permeances of 10,000 gpu (permeance [gpu] = permeability [Barrer]/thickness [µm] = 1000/0.1) could be prepared with today's technology, although in practice this will be challenging. Such an advanced membrane would reduce the cost and footprint of the membrane skids in a CO_2 capture plant by a factor of five compared to current membranes.

In addition to lab-scale membrane development, a number of membrane production runs were performed on MTR's commercial coating machine, using different membrane formulations. The purpose of this work was to scale up production of high-performance Polaris membranes developed in the lab, to ensure that large quantities of membrane could be made reproducibly. Table 1 shows pure-gas performance of membrane stamps taken at different locations of a sample Polaris membrane roll made on MTR's commercial coater. The roll was 40 inches wide and 120 feet long. The performance results indicate that the membrane properties were uniform over the length and width of the roll. The average CO_2 permeance was 1,300 gpu and the average CO_2/N_2 selectivity was 54 for the roll, consistent with stamps prepared in the lab using this batch of membrane materials.

Table 1. Pure-Gas Separation Performance for Polaris™ Production Run 030209. Measurements at room temperature (23°C) with 50 psig feed pressure and 0 psig permeate pressure.

Sample Position Along Roll Length	Sample Position Along Roll Width	Permeance (gpu)				Selectivity		
		N_2	CH_4	H_2	CO_2	CO_2/CH_4	CO_2/H_2	CO_2/N_2
50 ft	Left	25	83	110	1,330	16	12	54
	Middle	23	78	100	1,310	17	13	57
	Right	24	77	100	1,240	16	12	52
120 ft	Left	26	86	120	1,370	16	12	52
	Middle	26	84	110	1,290	15	11	49
	Right	21	71	90	1,270	18	14	60
Average		24	80	110	1,300	16	12	54

The membrane described in Table 1 was used to prepare several Polaris modules that were tested in the field with a flue gas slipstream at the Cholla power plant. The membranes showed stable performance over the test period; these results are discussed in detail in Section 3.

During this project, we performed extensive stability studies of Polaris membranes under simulated flue gas conditions. Figure 9 shows the time-dependent mixed-gas performance of Polaris membranes in three environments: a mixture containing 18% CO_2 in nitrogen, and the same mixture with SO_2 concentrations of 100 ppm and 1,000 ppm, respectively. There are two important things to note from this plot.

- First, the mixed-gas CO_2/N_2 selectivity (20-25) is lower than pure-gas values (50-55). This result is primarily due to the higher temperature in the mixed-gas experiments (50°C versus 25°C). Higher temperature reduces CO_2/N_2 selectivity in polymer membranes because CO_2 solubility decreases faster than that of N_2 with increasing temperature.
- Second, the CO_2 permeance and CO_2/N_2 selectivity of the Polaris membrane are stable over a period of 60 days, including over 40 days of permeating 100 ppm of SO_2 and 8 days of permeating 1,000 ppm SO_2. This result confirms that Polaris is a robust membrane not adversely affected by exposure to SO_2 in flue gas.

In addition, the experiments described in Figure 9 also showed that the membranes have SO_2 permeances approximately twice that of CO_2. Therefore, a membrane process that operates at 90% CO_2 capture, will co-capture over 95% of the SO_2 in the flue gas. A study of the potential benefit of co-capturing SO_2 with CO_2 is a worthwhile exercise, but beyond the scope of this report.

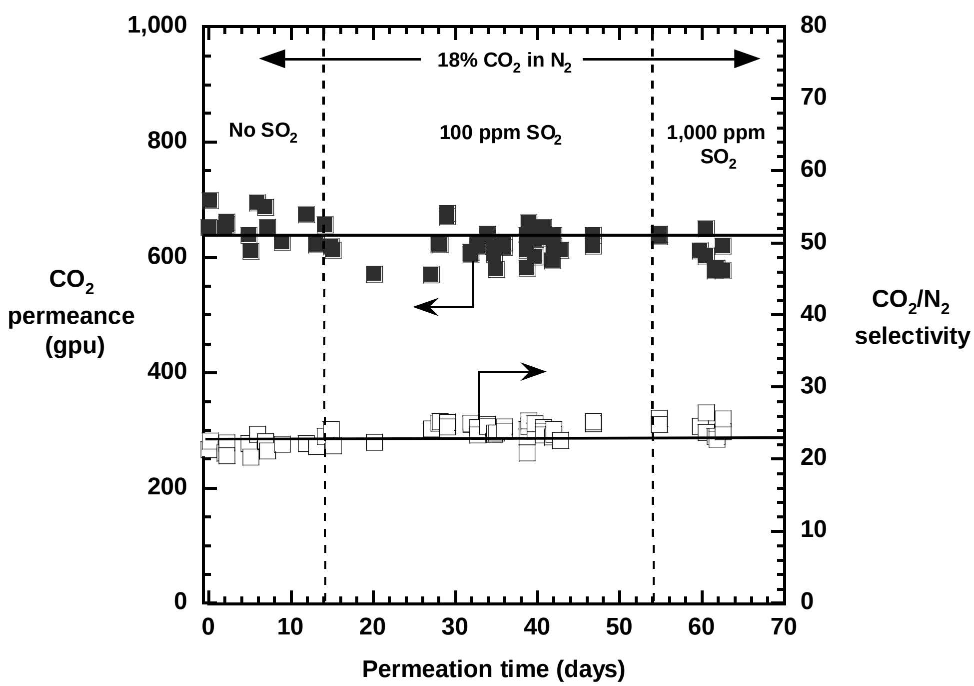


Figure 9. Time dependence of CO_2 permeance and CO_2/N_2 selectivity of Polaris™ membranes during continuous testing with simulated flue gas mixtures containing 0, 100, and 1,000 ppm SO_2. Temperature = 50°C; feed pressure = 50 psig.

Countercurrent/Sweep Module Development

In addition to membranes with high CO_2 permeance, a key innovation that makes capture of CO_2 in flue gas feasible with membranes is the use of combustion air as a sweep gas to generate driving force for separation. To utilize air for this purpose requires the development of countercurrent/sweep modules. Modification of a conventional spiral-wound module for use as the simplest possible counter-flow membrane contactor is illustrated in Figure 10. This figure shows an exploded view of a single membrane envelope. Two simple changes are required to achieve a countercurrent effect. First, the permeate collection pipe is closed in the middle, forming two separate compartments. Second, during module fabrication, additional glue lines are applied to direct gas flow in the permeate channel. As shown in Figure 10 (b), these modifications allow the permeate channel to be swept with a sweep gas and the module to operate in a countercurrent mode. Permeate gas flows countercurrent to the feed gas flow.

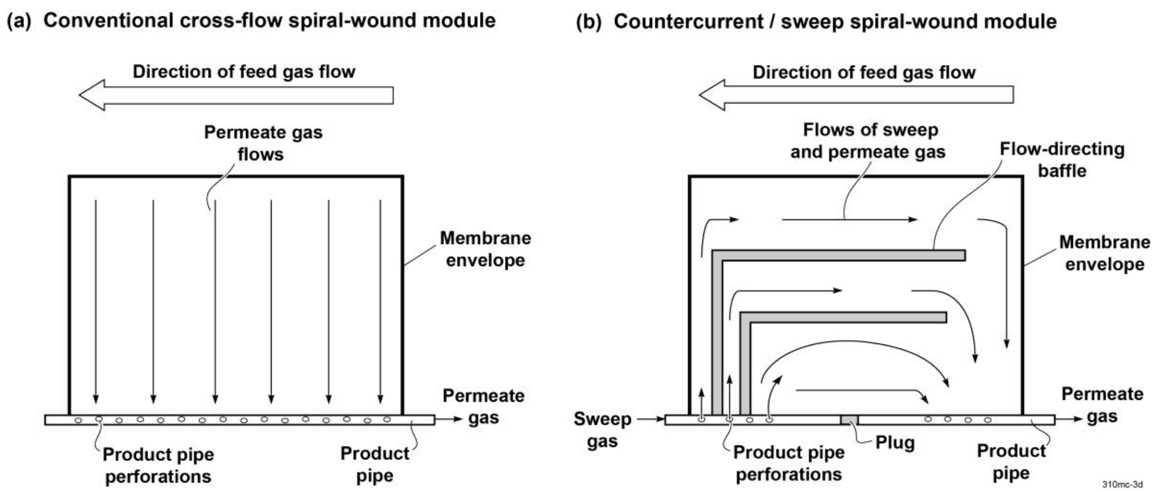


Figure 10. Unwound view of the membrane envelope for two types of spiral-wound modules. The flow pattern in the conventional module (a) is cross-flow, whereas the modified module (b) accepts a sweep gas on the permeate side and operates in a partial countercurrent pattern. Multiple spiral-wound modules of the sweep design can be housed in a single pressure vessel; the sweep gas flows from module to module through the permeate pipe connections.

In our previous DOE NETL project (DE-NT43085), Polaris membrane-based cross-flow and sweep modules with low feed side pressure drops were successfully developed. In this project, significant efforts were devoted to sweep flow channel optimization for countercurrent/sweep modules, in an attempt to:

- Reduce pressure drops in the sweep flow channels
- Maximize countercurrent flow for sweep module designs.

Pressure drop within module channels is undesirable because it increases the energy required to move gas through the system, and it reduces the driving force for permeation. The pressure drop in feed and sweep channels of a membrane module is caused by resistance to flow through the spacer materials that create the flow channels in a module. These spacers are porous media, and flow through these materials is governed by the Forchheimer equation:

$$\frac{dP}{dx} = \eta \frac{U}{k_1} + \rho \frac{U^2}{k_2} \tag{1}$$

where ρ is the density of the fluid, U is the superficial velocity of the fluid, k_1 is the Forchheimer viscous coefficient, k_2 is the Forchheimer inertial coefficient, and η is an empirical constant. The Forchheimer coefficients, k_1 and k_2, are determined by the spacer geometry and porosity. For a gas stream with a given flow rate, its superficial velocity in a membrane envelope is determined by the height of the flow channel (thickness of the spacer that creates the flow channel). Therefore, pressure drop can be reduced by increasing the height of the flow channel to lower the superficial velocity, or increasing the open space in the channel to increase the k_1 and k_2 values. These approaches must be balanced by the desire for high membrane packing density (increasing channel height reduces packing density) and the required membrane mechanical support (more open spacers provide less support for the membrane). In this project, we screened a number of spacers, and prepared numerous sweep modules using spacers with promising properties.

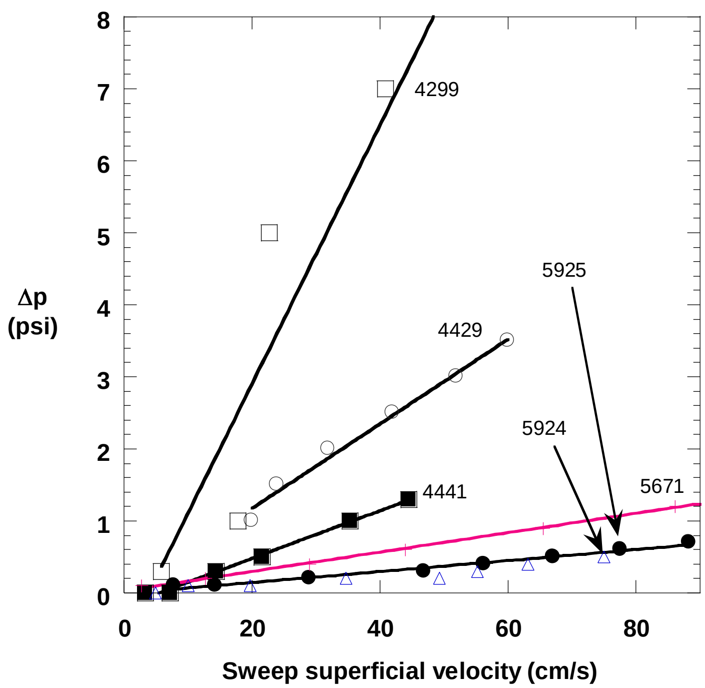


Figure 11. Module sweep-side pressure drop as a function of the sweep gas superficial velocity. Tests were conducted at 22°C with either 35 psia (modules 4299 and 4429) or 40 psia (modules 4441, 5671, 5924 and 5925) of nitrogen on the feed side of the module. Nitrogen was used as the sweep gas and exited the module at atmospheric pressure.

Figure 11 shows the sweep-side pressure drop as a function of sweep gas superficial velocity for the modules prepared in this project. The module with the highest serial number was made most recently. The modules prepared towards the end of this project (5924 and 5925) showed significantly lower pressure drop, compared to earlier sweep modules (4299 and 4429). This was achieved by using optimized spacer materials and flow channel configurations. The superficial velocity in the sweep modules for a full-scale capture system using parallel module design is expected to be in the range of 150-250 cm/s, higher than what can be achieved using our lab-scale module testing apparatus. Based on the results shown in Figure 9, the extrapolated pressure drops of modules 5924 and 5925 would be approximately 2 psi in a full-scale operation. These improvements are critical to the success of the membrane capture approach because each additional 1 psi of pressure drop in a full-scale, 550 MW_e system amounts to 2-3 MW_e of additional blower energy that must be supplied.

As pressure drop is reduced in a module, one concern is that such an open flow channel may result in poor mixing, and boundary layers that limit CO_2 transport will develop. Figure 12 compares the effectiveness of sweep operation in modules 4299, 5671, 5924 and 5925 for a simulated flue gas feed. To make this comparison, we calculated the theoretical CO_2 flux for each module estimated by computer simulation (ChemCad 5.6) and the resulting pure-gas module properties. These theoretical CO_2 fluxes were then compared to the actual measured values. The ideality of sweep module performance was calculated as the percentage of measured CO_2 flux relative to the theoretical maximum CO_2 flux. Figure 12 shows that permeate CO_2 fluxes of modules 5924 and 5925 are slightly lower than that of module 5671, but much higher than 4299. This result indicates that in addition to reducing sweep-side pressure drop, we have maintained, or even improved, the sweep performance in newer modules.

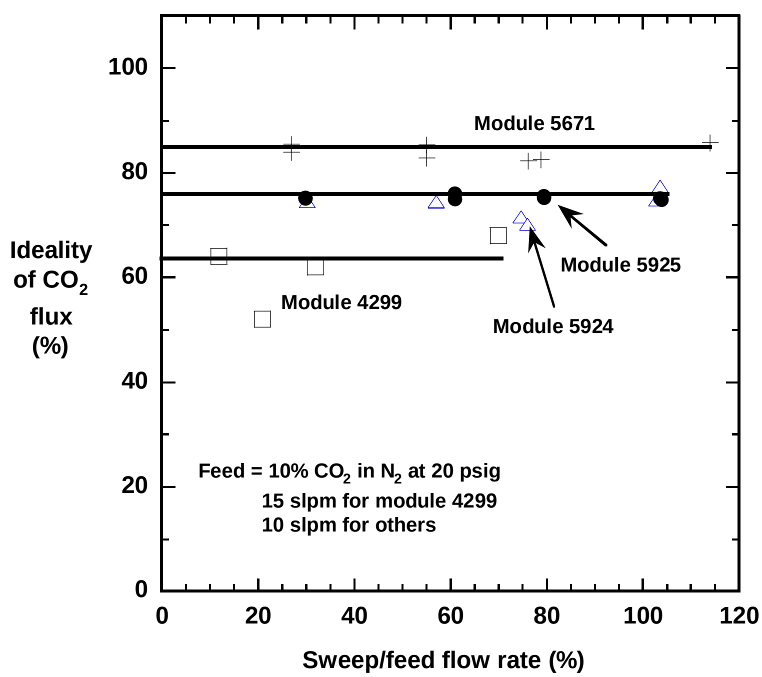


Figure 12. The ideality of CO_2 flux through modules 4299, 5671, 5924 and 5925 as a function of sweep-to-feed flow rate ratio. Temperature = 23°C.

In addition to pressure drop, another key module issue is stability in a flue gas environment. Water, oxygen, SO_2 and other minor contaminants have the potential to react with or otherwise damage materials used in membrane module construction. To investigate materials stability, tests of module components were conducted in a simulated flue gas environment (1,000 ppm SO_2, 18% CO_2, and liquid water at 50°C and atmospheric pressure). After exposure to simulated flue gas for over two months, all of the module components including glues, feed and permeate spacers, and support membranes maintained their functionality.

The module development work during this project allowed us to prepare modules with a combination of high performance and stability for the field tests at the Cholla power plant. The key challenge remaining is to further reduce module pressure drop without sacrificing transport performance. To date, we have evaluated a large number of spacer materials, and are currently using those with the highest Forchheimer coefficient k_1 and k_2 values (lowest resistance to mass transfer), available on the market. If we use a module configuration common for current industrial-scale gas separation membrane systems, four to six spiral-wound modules would be connected in series. This means the overall pressure drops across each membrane separation step for this application will be in the range of 1-2 psi for the feed side, and 4-8 psi for the sweep side, depending on the system design. These values, particularly for the sweep side, are too high for use in a large flue gas CO_2 capture application. To further reduce the sweep side pressure drops, a parallel flow pattern with minimal change in flow direction is required.

3. CHOLLA FIELD DEMONSTRATION

Field demonstration is the most effective way to evaluate the potential of membrane technology for post-combustion CO_2 capture. Working with Arizona Public Service (APS), MTR conducted a field demonstration during this project, at the APS Cholla coal-fired plant. The Cholla field test was the focus of this project, because the majority of the U.S. power generation installed base uses coal as a fuel. A system was built as part of our previous DOE NETL project (DE-NT43085) to enrich CO_2 from a flue gas slipstream. This test gave the project participants a chance to "pre-test" the membranes in conjunction with an actual flue gas feed, at a smaller scale.

Background

In May 2009, a membrane system was installed at APS's Red Hawk natural gas-fired power plant located west of Phoenix. The Red Hawk plant is comprised of two identical 530-megawatt natural gas-fueled combined-cycle units, and began operation in mid-2002.

The Red Hawk system used a commercial-scale (8-inch-diameter; 20 m^2 membrane area) Polaris module. Figure 13 shows a picture of the MTR system at Red Hawk. This system used a simple one-stage design to concentrate CO_2 from 3-4% in the flue gas to 20-30% in the gas going to a site testing the use of CO_2-fed algae for biofuel production.

Figure 13. Photograph of the MTR Polaris^{TM} membrane skid for capturing CO_2 from flue gas at the APS Red Hawk natural gas-fired power plant.

The Red Hawk system started operation in July 2009. The feed to the membrane system was natural gas power plant flue gas at 50°C, containing 3.6% CO_2, 13.4% O_2, 1.8% H_2O and 81.2% N_2. The membrane system enriched CO_2 from 3.6% in the feed to about 20% in the permeate.

Figure 14 shows the CO_2 permeance of the Red Hawk module as a function of time. The initial data point from April 2008 is a pure-gas quality control measurement taken at MTR after module fabrication. The module was then stored at APS for about 14 months prior to system startup. The data points from July and September 2009 are mixed-gas values obtained while running the module with raw flue gas from the power plant. For all cases, the module CO_2 permeance was around 600 gpu. This result indicates that the module was stable after 17 months of storage and operation.

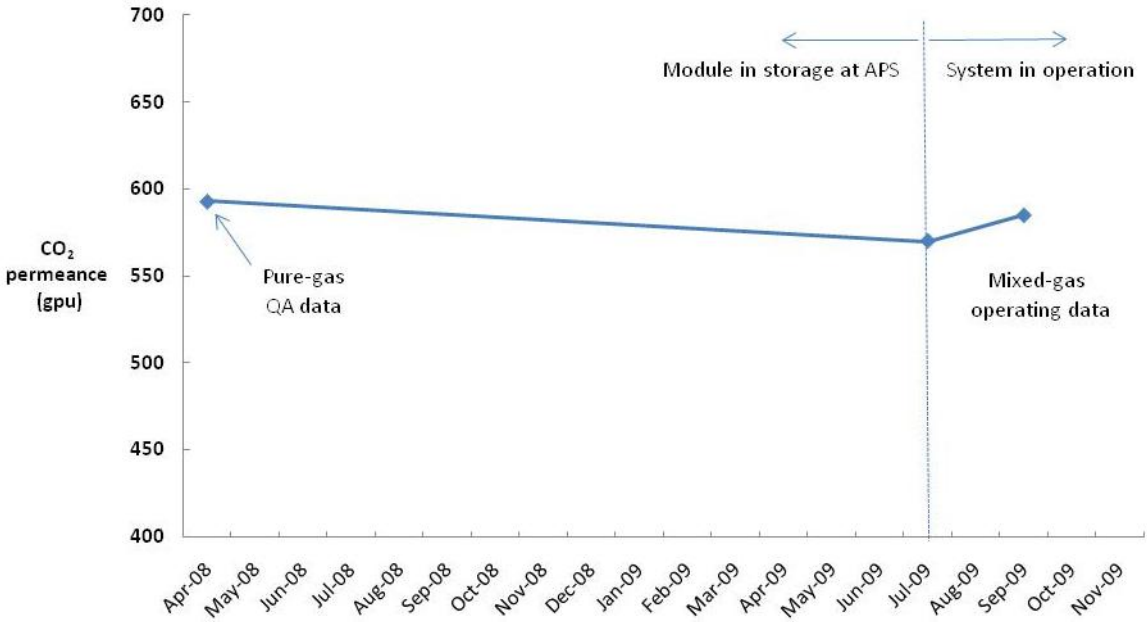


Figure 14. CO_2 permeance of the Polaris^{TM} module installed at the APS Red Hawk power plant. The initial pure-gas permeance from April 2008 was measured at MTR after module fabrication. The data points from July and September 2009 are field measurements of the module treating raw stack gas.

While the CO_2 flux through the Red Hawk module was consistent with our design estimates, the permeate CO_2 purity was lower than expected (20% versus 30%). This result was caused by a number of operational factors that can be summarized as follows:

- The feed flow rate was lower than expected due to an undersized slipstream feed line; this increased the module stage-cut and lowered the driving force for CO_2 permeation.
- The flue gas contained a higher O_2 concentration (13.4% O_2) than expected, possibly due to leaks in the slipstream feed line. Compared to nitrogen, oxygen permeates Polaris membranes more easily. More oxygen permeating through the membrane results in lower CO_2 permeate purity.
- The flue gas contained lower CO_2 concentration in the feed than expected (3.6% compared to 4.2%), again possibly due to line leaks. This is a relatively significant percentage difference in feed CO_2 content and reduced the CO_2 permeation driving force.
- High operating temperature and significant O_2 content resulted in relatively low membrane selectivity. Our measurements suggested a module CO_2/air selectivity of about 15 at 50°C.

These issues illustrate the importance of obtaining accurate process information when designing a membrane system and ensuring that the slipstream gas accurately represents the process gas. Overall, the performance variations from the original design were not critical to this demonstration, for which the main goal was to evaluate membrane stability.

Following startup, the Red Hawk system ran intermittently through 2009, while APS made adjustments to control systems outside the membrane skid. Due to management changes at APS in early 2010, APS decided to discontinue testing with the Red Hawk system. Nevertheless, the operational experience in a power plant environment gained during this field test provided positive information on the membrane module performance and lifetime, and helped clarify power plant modification and integration issues for the Cholla demonstration.

Cholla Field Test

Treating a coal-fired flue gas slipstream from the APS Cholla power plant was the focus of this project. The Cholla power plant is located in northeastern Arizona, near Holbrook, and has a total capacity of 995 MW_e. The plant is fueled by coal from McKinley Mine in New Mexico. The objectives of the Cholla test were to:

- Demonstrate performance of commercial-sized modules with real coal-fired flue gas;
- Determine pretreatment requirements (ash removal, water handling);
- Demonstrate air sweep operation in commercial-sized modules, and
- Obtain experience in operating rotating equipment with real flue gas.

System Installation and Operation at Cholla. Construction of the membrane system for use at Cholla was completed at the end of 2009. The skid was then delivered to APS and installed in a bay under the unit 3 boiler electrostatic precipitator (ESP), which was close to the flue gas slipstream port. As shown in Figure 15, post-FGD flue gas was taken from the top of the FGD unit and delivered to the membrane system by an 8-inch CPVC pipe. The exhaust gas from the membrane system, including CO_2-enriched permeate, membrane-treated flue gas, and sweep air was returned to the inlet of the ID fan in front of the FGD.

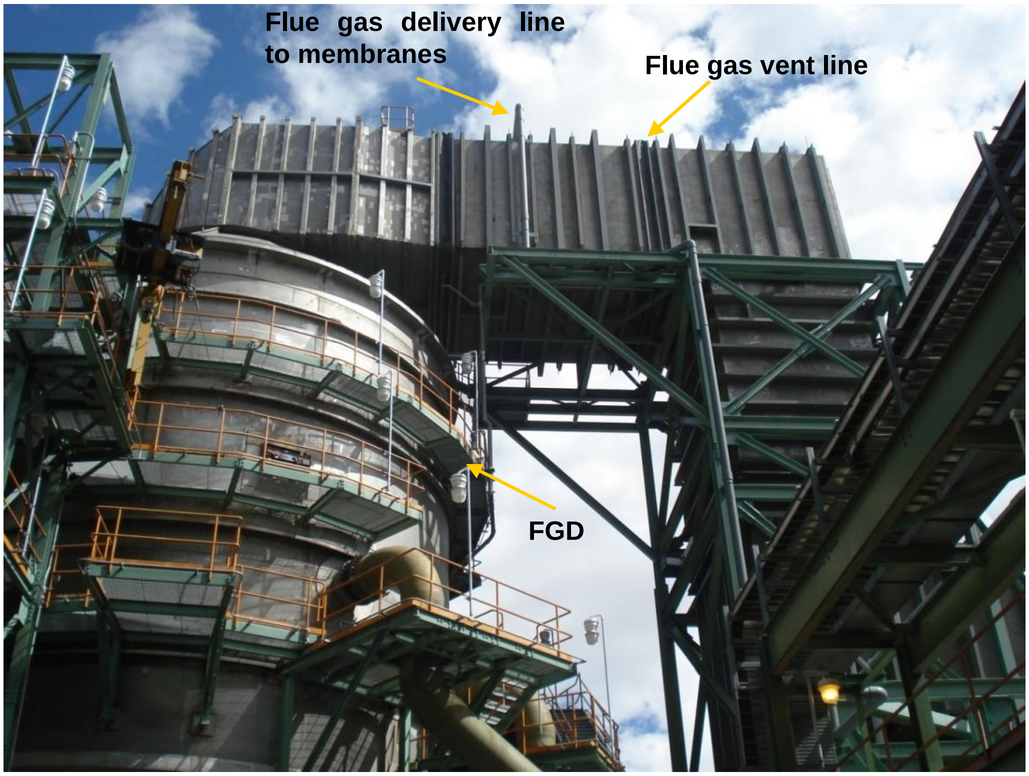


Figure 15. Photo of post-FGD flue gas line at Cholla, showing delivery line installed for flue gas delivery to the MTR membrane test skid.

Figure 16 shows a picture of the MTR membrane test skid at Cholla during installation. The skid contained two module pressure vessels – one designed to hold cross-flow modules and the other to accommodate countercurrent/sweep modules. Each pressure vessel could house up to four 8-inch-diameter Polaris membrane modules. The skid was fitted with an array of flow, pressure, temperature, and gas composition analyzers for system performance monitoring.

The 8-inch diameter CPVC feed line that delivered flue gas to the membrane system from the outlet of the Cholla FGD unit is shown at the top of the Figure 16 photograph. This flue gas feed was sent to a compressor that provided feed to the membranes at pressures from just above ambient pressure to 80 psig. The feed gas was sent first to cross-flow modules, where a vacuum on the permeate provided driving force for CO_2 capture. The residue gas from these modules was then sent to sweep modules, where an air compressor provided the sweep gas to generate driving force for CO_2 transport. After measurements, the residue and permeates from the system were combined and returned to the front of the Cholla FGD unit. This membrane system design is shown in Figure 17.

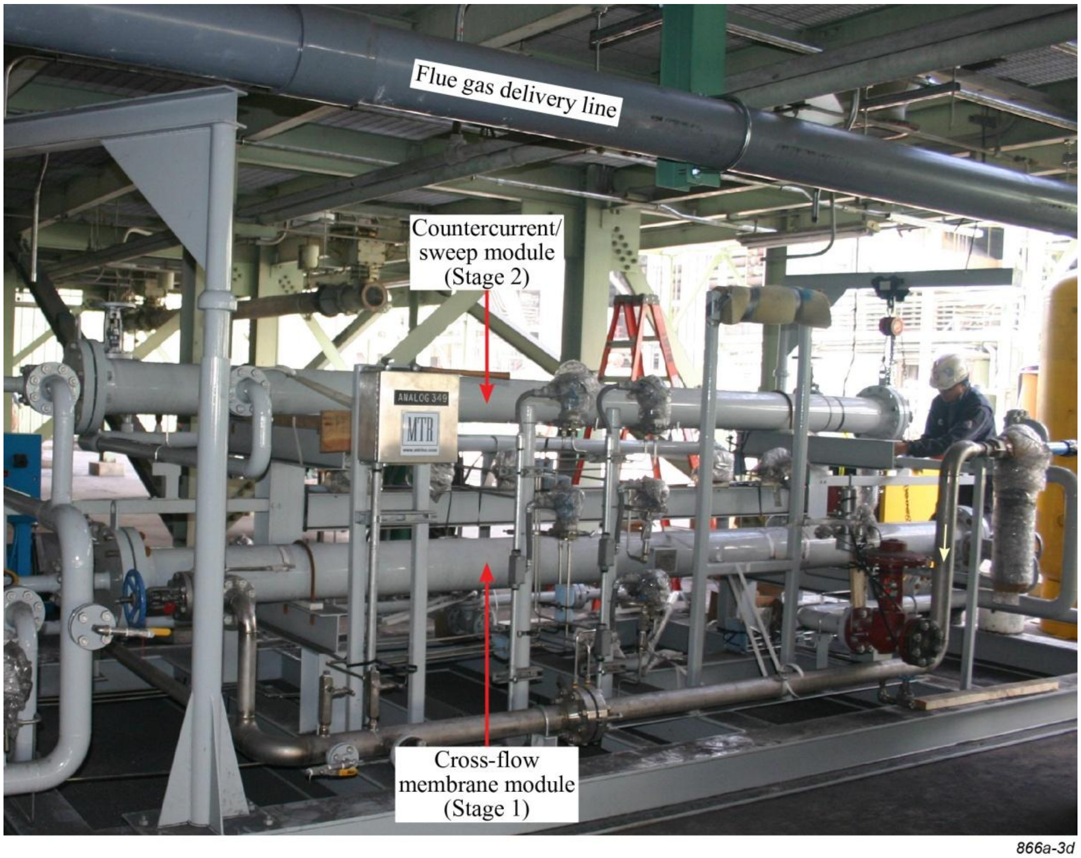


Figure 16. A picture of the MTR membrane test skid at Cholla during installation.

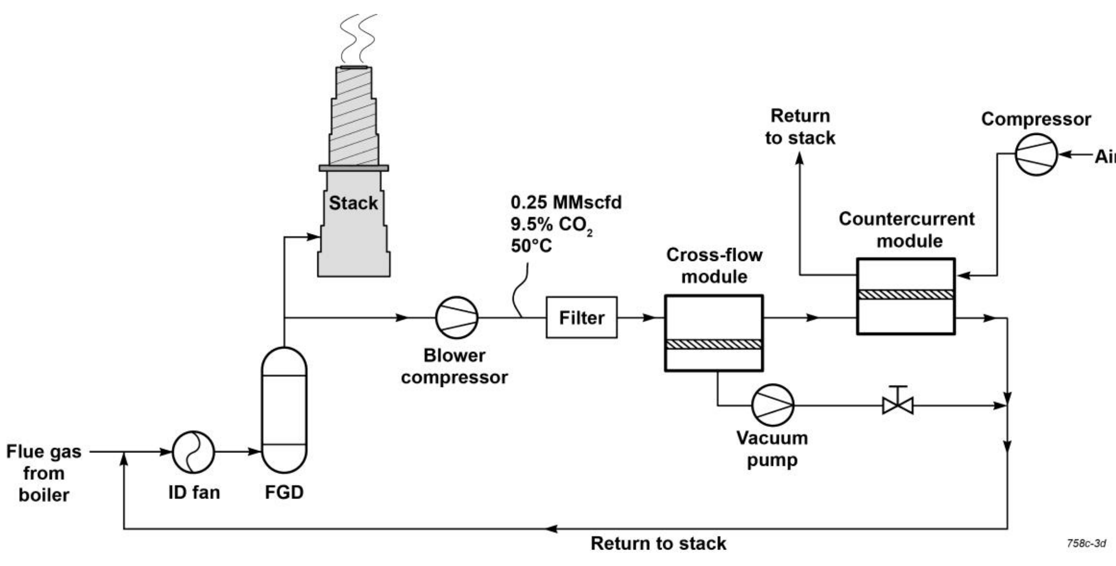


Figure 17. Flow diagram of the membrane process for the Cholla field demonstration.

Compared to a full scale design, the main difference for the slipstream test system is that the sweep outlet gas is returned to the stack rather than being recycled to the boiler. The small size of the slipstream system compared to the size of the Cholla boilers precluded a meaningful study of boiler recycle. Figure 18 shows a photograph of the 8-inch-diameter Polaris membrane modules being installed at Cholla. These modules included a number of new features that were tested, including low-cost end caps and novel spacer configurations.

Figure 18. A picture of the 8-inch diameter Polaris modules being installed in the membrane test system at Cholla.

The Cholla system began operation in mid-April 2010, and ran for approximately three months. During this period, the system was in operation about 45% of the time. Both scheduled and unscheduled shutdowns accounted for the remaining time. Figure 19 shows the run time history of the membrane system. The scheduled shutdowns involved modifications to the system to allow for better control of system variables (such as feed pressure and flow rates) and upgrades to the analytical equipment to allow more remote monitoring. The unscheduled shutdowns centered around rotating equipment problems, primarily with the feed compressor and vacuum pump, resulting from water and corrosion issues.

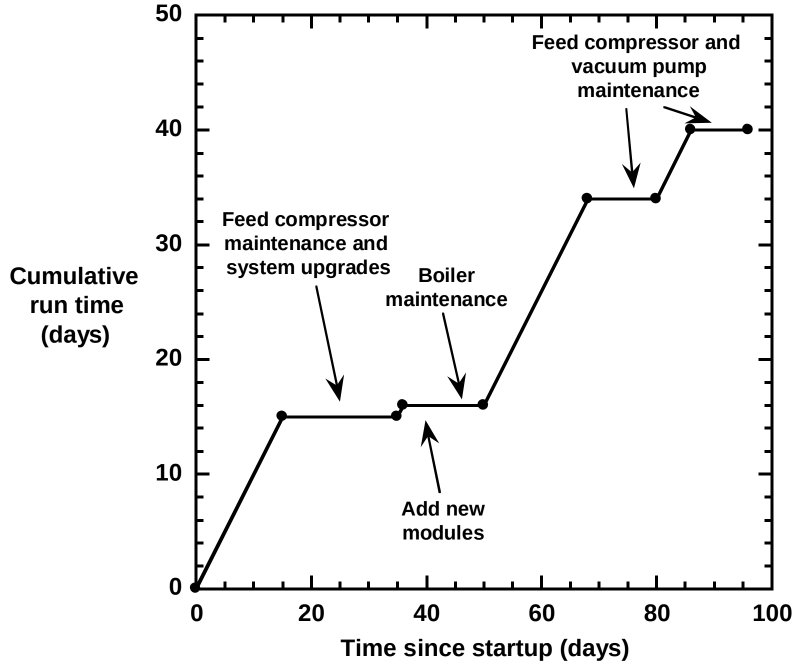


Figure 19. Cumulative run time history of the Cholla skid.

We found that any non-stainless steel equipment component was susceptible to corrosion and failure. This included even electrical panels in the feed compressor that are nominally isolated from process gases. Figure 20 shows pictures of a vacuum pump motor after three months of operation; its shaft (a carbon steel shaft, mistakenly supplied by the manufacturer) was completely dissolved by the permeate gas. In August, we shut down the system to repair the pump, replace the feed compressor and add an appropriate flue gas pretreatment section.

Figure 20. Pictures of the first vacuum pump motor installed in the Cholla membrane unit, showing the remaining shaft stub after the shaft was dissolved by the permeate gas.

In the original project scope, conditioning of the flue gas, including water knockout, was to be handled by APS. However, turnover on the APS team during the construction/installation part of this project changed the work scope. As a result, raw post-FGD flue gas was delivered to our membrane skid without any pretreatment. The acidic liquid water in the flue gas caused equipment corrosion issues that accounted for most of the system downtime during the field test. Future testing will require careful materials selection and pretreatment consideration.

System Performance at Cholla. Test results from the Cholla system are shown in Figures 21 and 22. In these figures, the test period has been divided into four segments corresponding to the following different run conditions:

- In the first period (I), the system started operation with two cross-flow and two sweep modules (total system capacity is four of each module type). We intentionally undersized the membrane area for start-up, to limit the number of modules that might be damaged by unexpected start-up upsets. In this configuration, the system captured about 0.6 ton CO_2 per day from the post-FGD flue gas. The CO_2 concentration in the flue gas was reduced from 9.5% in the feed to about 5% in the residue, and a CO_2-enriched permeate containing 42-45% CO_2 was produced. Other than some minor fluctuations due to temperature and flow rate variations, the system exhibited stable performance over this period. The two cross-flow modules were removed at the end of this period and showed good performance in a post-test evaluation (the results are described below in Table 2).

- In the second period (II), we increased the system CO_2 capture capacity by increasing the number of cross-flow membrane modules in the first stage to three. The system then removed about 0.9 ton CO_2 per day, which amounts to 80-85% of the CO_2 in the flue gas inlet. The treated flue gas leaving the membrane skid contained only around 2% CO_2. At the end of this period, we replaced the countercurrent/sweep modules and conducted post-test analysis on the two modules removed (see Table 2 discussion).

- In the third period (III), we lowered the system feed pressure from 70 psig to 35 psig. This feed pressure is similar to that identified from process design studies as the optimum condition to minimize parasitic energy loss. As expected, this decrease in feed pressure reduced the amount of CO_2 removed. To compensate, we planned to add additional modules to operate the system at 90% CO_2 capture. However, we were unable to do testing with maximum module capacity (and 90% CO_2 capture) during this project because APS decided not to continue testing at Cholla.

- In the fourth period (IV), the system was running without the vacuum pump due to pump shaft failure. The driving force for gas permeation in cross-flow modules was only provided by the feed compression. As a result, the feed-to-permeate pressure ratio was significantly reduced, resulting in less CO_2 capture by the cross-flow modules and lower CO_2 concentration in the CO_2-enriched permeate stream.

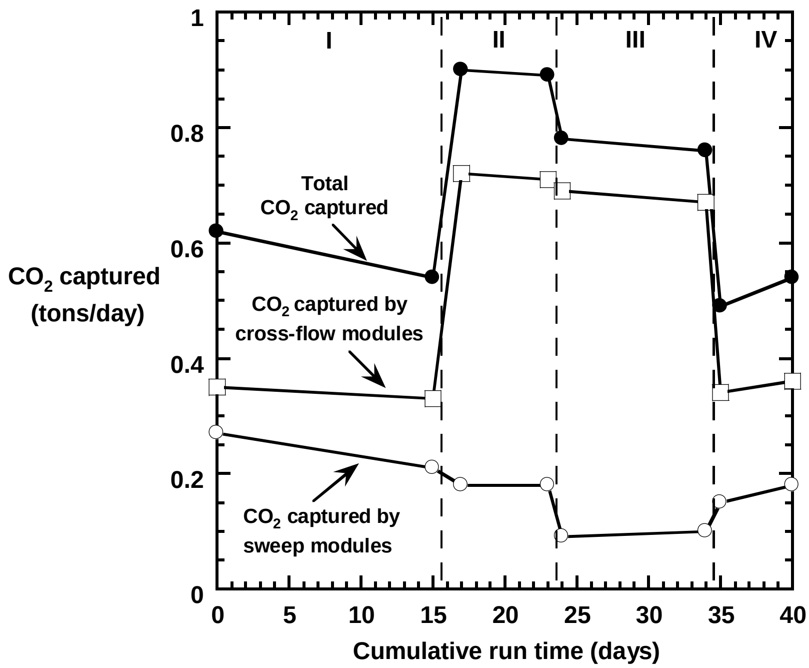


Figure 21. Amount of CO_2 captured by the two-step Cholla membrane skid, the cross-flow modules only, and the countercurrent/sweep modules only.

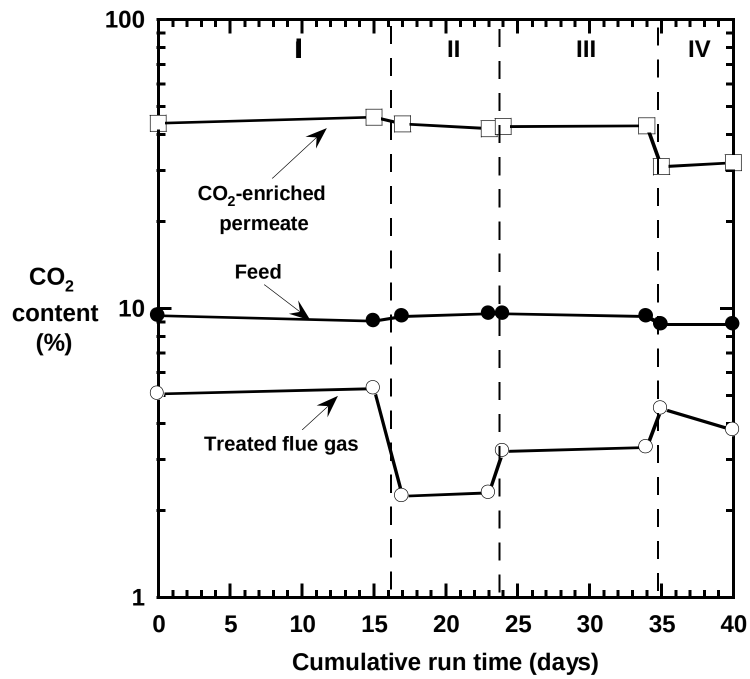


Figure 22. CO_2 concentrations of post-FGD flue gas (feed), membrane-treated flue gas (residue), and the CO_2-enriched permeate measured for the Cholla skid.

As shown in Figure 22, throughout the first three periods, the CO_2 concentration in the first-step permeate (vacuum) was 42-45%, which is nearly a five-fold enrichment over the feed gas (9.5%). This result is in line with predicted performance for the system operating conditions. In a larger system, where sweep recycle to the boiler is used to increase the feed CO_2 content to 18-20%, the permeate CO_2 content would be greater than 75% on a dry basis. The residue or treated flue gas stream contained between 2% and 5% CO_2, depending on the operating conditions described above. The amount of CO_2 in the treated gas stream was also consistent with predicted performance, and can be easily tuned by changing the feed pressure or membrane area (that is, the number of modules).

During the Cholla tests, modules were periodically removed from the system and returned to MTR for analysis. Table 2 shows the relative performance results measured at MTR for two of these modules before and after testing at Cholla. Within the uncertainty in the measurements, the module performance was unchanged after 45 days in the field treating harsh post-FGD flue gas (saturated with water, 50 ppm SO_2, residual particulates, and so forth).

Table 2. Module Performance Before and After Operation at Cholla

Module Number	Test Date	Room Temperature (°C)	Normalized CO_2 Permeance	Normalized CO_2/N_2 Selectivity
5839 (Cross-flow)	01/27/2010	18.5	100%	100%
	08/02/2010	21.8	110%	118%
5879 (Sweep)	02/24/2010	15.7	100%	100%
	08/03/2010	22.6	108%	96%

Figure 23 shows a picture of one of the modules installed at Cholla before and after 45 days of treating post-FGD flue gas. This cross-flow module was the first one in series and was directly exposed to the incoming flue gas. There is clearly discoloration on the annular surface of the module where the flue gas enters the feed side flow channels. Analysis of the module showed that most of the debris on the surface was rust, presumably from the corrosion of upstream system equipment. Replacement of non-stainless steel components on the feed compressor and gas cooling system should alleviate this problem.

(a) (b)

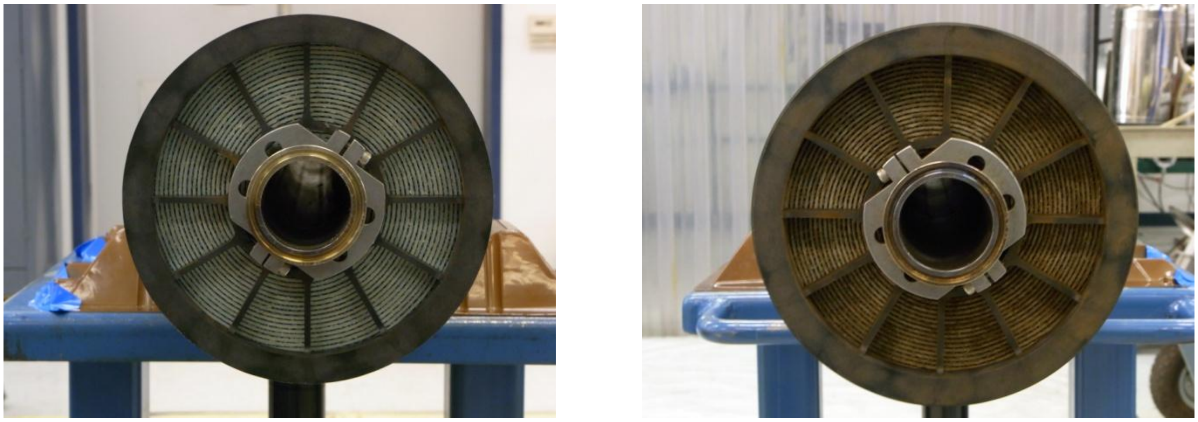


Figure 23. Photographs of a cross-flow module (a) before and (b) after 45 days of exposure to post-FGD flue gas at Cholla.

Countercurrent/Sweep Membrane Module Operation

Use of combustion air as the sweep gas to provide “free” driving force for CO_2 separation is a key enabling aspect of our membrane-based CO_2 capture process (U.S. Patent 7,964,020[11]). Therefore, one task of the demonstration was to confirm the effectiveness of MTR’s commercial-scale sweep modules operating with real flue gas. Figure 24 shows results obtained from the Cholla field test. Lab results (solid circles) and theoretical performance calculated by our process simulator (solid line) are also shown in the figure for comparison. Clearly, when the air sweep gas is turned on, there is a substantial increase in CO_2 flux compared to non-sweep operation. For example, the CO_2 flux through the modules increases by roughly five times as the sweep/feed flow ratio is increased from 0 to 50%. At higher sweep/feed flow ratios, the CO_2 flux begins to level off, as predicted by theory. Both field and lab results are slightly lower than the ideal CO_2 flux values (about 80% of ideality over the flow range tested). These results are an important confirmation that CO_2 capture with a sweep membrane design can work in practice.

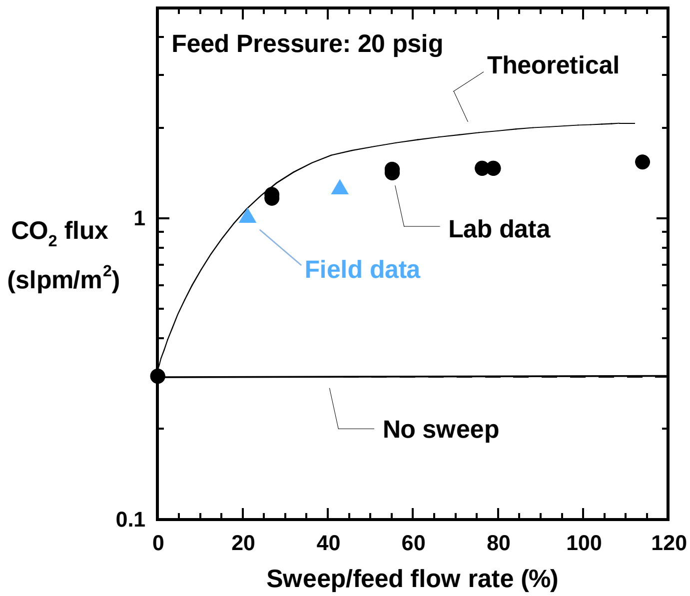


Figure 24. The influence of sweep flow rate on CO_2 flux through an MTR CO_2 capture Polaris membrane fabricated into countercurrent/sweep spiral-wound membrane modules.

System Upgrade and Future Field Demonstration

Based on the operational experience collected over three months of run time at Cholla, acidic water condensate in the flue gas slipstream pipeline flowing into the feed compressor was the main cause of system corrosion and shutdowns. Based on our analysis, the flue gas feed stream contained about 30 kg/hr of water with a pH of 2.0 – 2.5. Any carbon steel or other non-corrosion resistant material in contact with this condensate was highly susceptible to rust formation.

To address these corrosion issues, a number of changes were made to the membrane skid to allow future testing to proceed more smoothly. A pretreatment system was installed on the skid to cool the flue gas and condense the bulk of the process water. Redundant knockout drums and heat tracing were installed to prevent two-phase flow from entering the feed compressor. In addition, a new feed compressor with coated rotors rated for acid gas service was installed and the permeate vacuum pump was replaced with an appropriate model using a stainless steel shaft.

The new pretreatment system was ready for installation in November 2010. However, due to changes in the APS management team, APS decided to stop participation in this program and in our next project – a 1 MW_e field test of the technology. In early 2011, MTR began preparing to conduct additional testing of the 1 ton/day skid at the National Carbon Capture Center (NCCC) run by Southern Company. This future testing will help MTR obtain more extensive membrane lifetime and performance data that will be used to design the larger 1 MW_e equivalent membrane system.

In summary, the 3-month Cholla field test is, to our knowledge, the first ever test of commercial-scale membrane modules for CO_2 capture from real coal-fired power plant flue gas. During the test, all membrane modules showed stable performance. Membrane fouling by residual particulates – a major concern entering this project – was not a factor over the timescale of this demonstration. We also confirmed the effectiveness of using air as a sweep stream in commercial-scale countercurrent modules, which is a key to capturing CO_2 in a cost-effective manner. Most of the problems encountered during the Cholla test occurred because of corrosion problems with rotating equipment. New components and materials have been added to the 1 ton/day skid to alleviate this problem in future testing. The operational experience gained from the Cholla test will prove valuable in permitting future tests to run more smoothly.

4. PROCESS DESIGN CONSIDERATIONS

Introduction

The membrane process design issues associated with flue gas CO_2 capture were introduced in Section 1. The key points can be summarized as follows:

- Membrane processes require pressure driving force to separate gases. Generating this driving force, in the form of pressure ratio, is costly for flue gas treatment because of the large flow rate and low pressure of flue gas.
- Because the affordable pressure ratio for flue gas CO_2 capture is limited to 5 or perhaps 10 at most, high membrane CO_2/N_2 selectivity is not helpful. High CO_2 permeance is more useful to reduce the required membrane area.
- Because of the affordable pressure ratio limitation, a single-stage membrane process will not be able to give high purity (>95%) CO_2 at high CO_2 capture rates, regardless of the membrane properties. Multi-stage membranes and/or hybrid processes will be required to capture 90% of the flue gas CO_2 at high purity.

Initial MTR CO_2 Capture Membrane Design

To address these challenges, MTR initially proposed the CO_2 capture membrane design introduced in Figure 6, and reproduced in simplified form here in Figure 25.

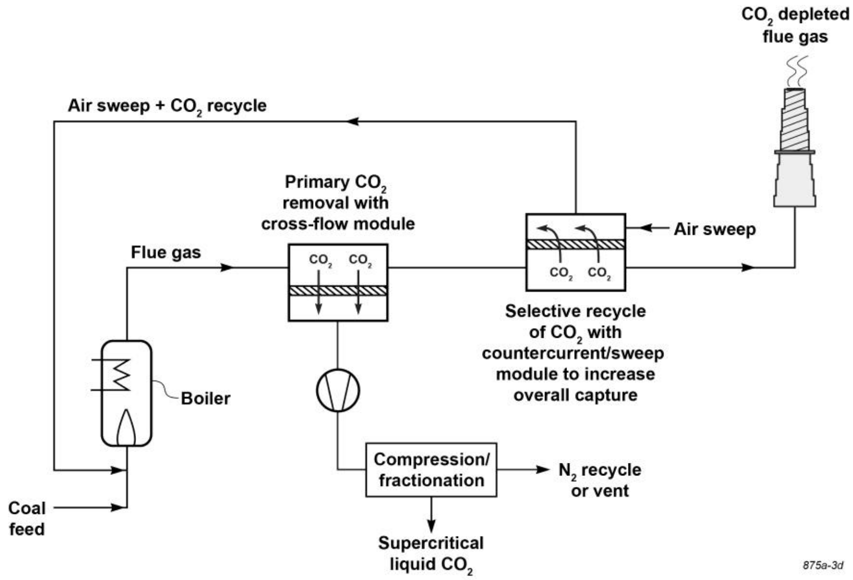


Figure 25. Simplified diagram of membrane-based selective recycle of CO_2 in series with a membrane CO_2 capture step as applied to combustion flue gas. See Figure 6 for additional details.

This design combines a conventional cross-flow membrane step with a novel countercurrent/sweep membrane step to remove 90% of the flue gas CO_2. The CO_2 captured by the membranes is then compressed and purified in a membrane-assisted refrigeration step to produce sequestration-ready liquid CO_2. The key aspect of the Figure 25 process design is the use of air in a countercurrent/sweep membrane unit that selectively recycles CO_2 not captured by the first membrane back to the boiler. Using air as a sweep stream generates partial pressure driving force without requiring costly and energy-intensive compressors or vacuum pumps. From the standpoint of CO_2 capture energy, there are two benefits to the selective recycle of CO_2 using a membrane unit:

1. By increasing the CO_2 content of the flue gas, selective CO_2 recycle reduces the minimum work (or energy) required to capture CO_2 using a membrane or any other capture technology, and
2. When operated in a serial manner, in which the membrane selective recycle step treats the CO_2-depleted off-gas of the capture step (as shown in Figure 25), the membrane selective recycle reduces the amount of CO_2 that must be removed in a single pass by the capture step. For example, the capture step may remove only 50% of the flue gas CO_2 and allow the membrane selective recycle step to re-circulate most of the remaining CO_2 so that 90% overall capture is achieved. By decreasing the amount of CO_2 that must be removed by the capture step in a single pass, the minimum work required to capture the flue gas CO_2 is reduced.

Minimum Energy of CO_2 Capture

Starting with a brief explanation of minimum energy calculations, a quantitative explanation of how selective CO_2 recycle using membrane sweep reduces the minimum energy of separation is given in the following section. This is followed by a general discussion of how sweep membrane units might be used in CO_2 capture schemes, with particular emphasis on hybrid designs in which a sweep membrane unit is used as a CO_2 pre-concentrator in conjunction with other CO_2 capture technologies. The membrane-only approach (highlighted in Figure 6) that is the base case for this report is analyzed separately in greater detail in Sections 5 and 6.

Recently, a number of researchers have investigated the minimum amount of energy (or work) required to separate CO_2 from flue gas and subsequently compress and liquefy the CO_2.[1,2] The thermodynamic minimum energy is obtained by calculating the difference between the Gibbs free energy of the initial stream to be separated and the Gibbs free energy of the streams produced after separation, compression and liquefaction. Herzog[12] and McGlashan[13] both use the ideal gas law to quantify the entropy effects; their results therefore represent the ideal minimum energy.

We analyzed the same simplified process used by Herzog and McGlashan (shown in Figure 26), but used a commercially available process simulator (ChemCad 5.6) to obtain the Gibbs free energy values of the streams, thereby utilizing the thermodynamic database incorporated in the simulator. Our results match very well with the results of Herzog and McGlashan, which means that non-idealities are not significant in the current process, most likely because the gas pressures are low.

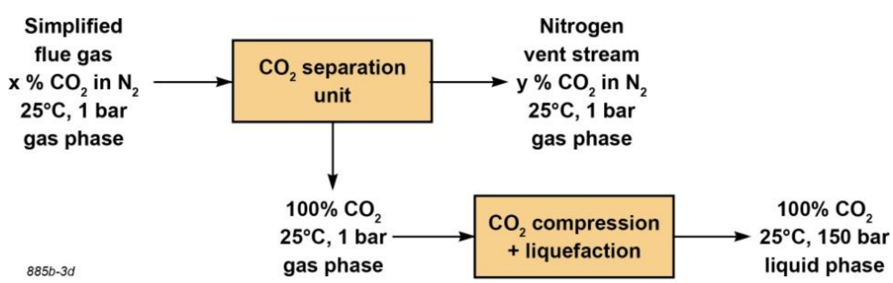


Figure 26. Simplified process schematic of CO_2 separation and compression/liquefaction for a simplified flue gas mixture of CO_2 and nitrogen. The CO_2 concentration in the feed (x%) depends on the type of flue gas. The CO_2 concentration in the nitrogen vent stream (y%) is a function of CO_2 removal and is zero at 100% CO_2 removal.

The results of our minimum energy analysis are presented in Figure 27, which gives the minimum energy required per ton of CO_2 captured as a function of the CO_2 content of the simplified flue gas stream (CO_2 plus nitrogen). The minimum energy is shown for three different levels of CO_2 capture. Also indicated is the minimum energy required just for CO_2 compression and liquefaction, which is independent of the initial CO_2 concentration.

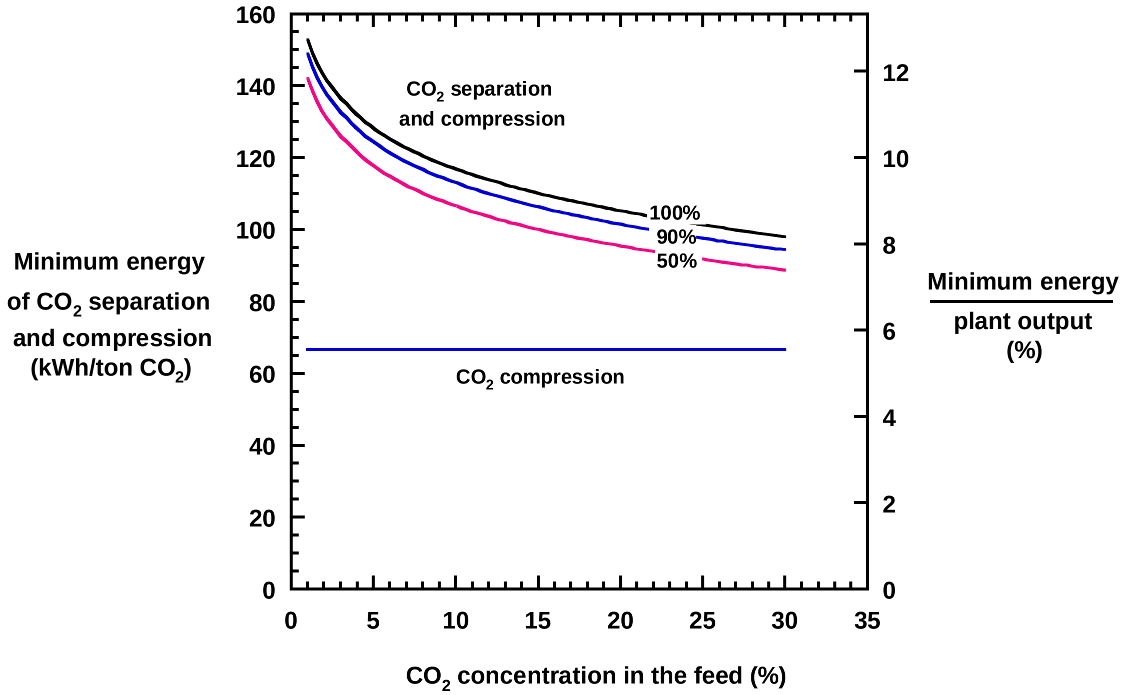


Figure 27	Minimum energy per ton of CO_2 captured as a function of CO_2 concentration in the flue gas. The ratio versus power plant output is calculated assuming that the energy expended is electric, not thermal. Calculations also assume a power plant produces 1,200 kWh per ton of CO_2 produced (10,000 ton CO_2/day for 500 MW_e). Minimum energies were calculated with a process simulator using the SRK equation of state.

The analysis shows that:

1. The minimum energy required for 90% CO_2 capture in a coal-fired power plant (feed CO_2 concentration around 13%) is about 110 kWh/ton CO_2 or about 9% of gross power plant output (assuming all energy expended is electric, not thermal). Approximately 43 kWh/ton CO_2 (3.3% of plant output) are required to separate 90% of the CO_2 from the flue gas feed and 67 kWh/ton CO_2 (5.5%) are required to compress the captured CO_2 to 150 bar.
2. The minimum energy required for CO_2 compression plus liquefaction is more than half of the total minimum energy when the feed CO_2 concentration is over 5%. However, because the actual compression/liquefaction step will be more efficient than the actual separation step, we expect an actual process will consume more energy in the separation step (discussed below).
3. The minimum energy required decreases if the percentage of CO_2 capture is reduced, albeit not very significantly. In practice, however, we expect that the separation step will operate at higher efficiency if the percentage of CO_2 capture is lower.

Impact of CO_2 Recycle Using a Membrane Sweep Unit on Minimum Energy

Figure 28 illustrates the impact of selective CO_2 recycle using a sweep membrane unit on the minimum energy required for the CO_2 separation step. The data assumptions in Figure 28 are the same as those in Figure 27, except that the CO_2 compression-liquefaction energy has been omitted, because this step is independent of the concentration of CO_2 in the feed to the separation step.

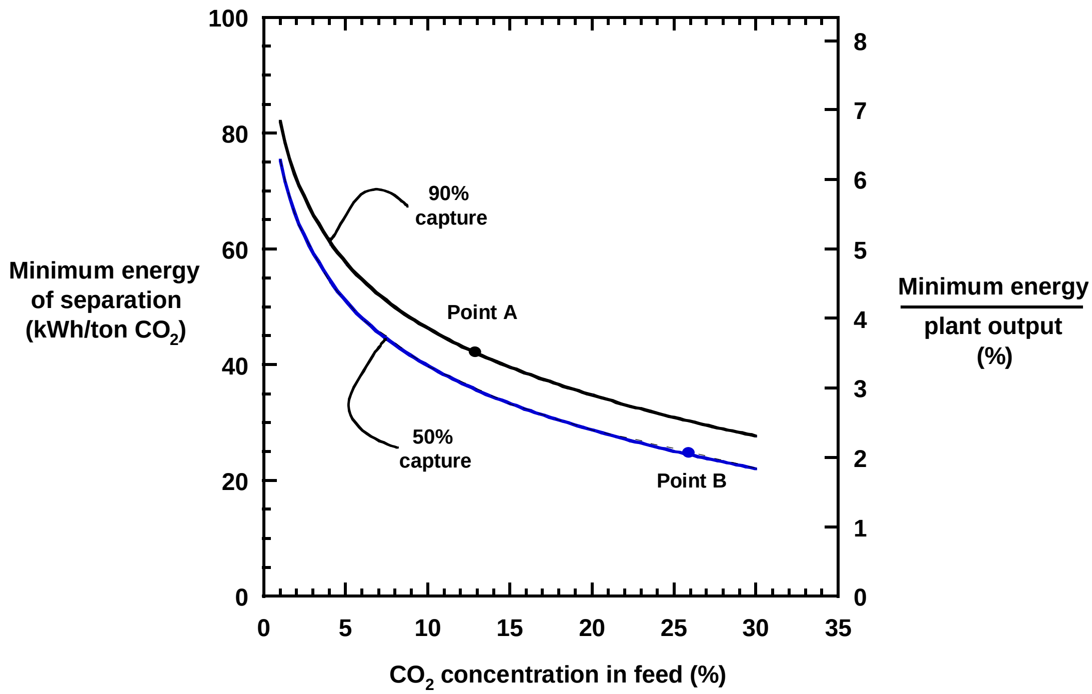


Figure 28. Minimum energy of separation as a function of CO_2 content in the feed stream.

Point A in the figure represents the minimum energy of 90% CO_2 separation for coal flue gas containing 13% CO_2. Point B in the figure is the minimum energy for the case where a sweep membrane unit has been used to concentrate the flue gas to 26% CO_2 and reduce the capture step requirement to 50% CO_2 removal (90% overall capture is still achieved because the sweep membrane unit recycles most of the remaining CO_2). For this example, the minimum energy of CO_2 separation is reduced from 43 kWh/ton CO_2 to 25 kWh/ton CO_2. This is a reduction of 42% in the minimum energy of separation. Most of this reduction in minimum energy is due to increasing the concentration of CO_2 in the flue gas (moving to the right on the x-axis in Figure 28); a small portion of the reduction is due to the decrease in the removal requirement for the capture step (moving from the 90% to the 50% capture curve on the same figure).

The actual energy of CO_2 capture based on currently available separation technologies is much higher than the minimum energy. For example, Herzog[12] estimates that the actual energy of 90% CO_2 separation for a state-of-the-art amine capture process treating coal flue gas is five times the minimum energy or about 16% of the power plant output. Based on the Figure 28 analysis, increasing the CO_2 concentration by using a sweep membrane unit can potentially reduce the actual energy use of any current separation process by 42%, to about 10% of power plant output. This energy savings (10% of plant output with membrane, versus 16% without) would be substantial, amounting to 30 MW_e at a 500 MW_e power plant.

Alternative Process Designs

One of the tasks in this project was to evaluate potential advantages of alternative membrane process designs. The capture process shown in Figures 6 and 25 is an all-membrane design that takes advantage of a sweep membrane to efficiently remove CO_2 from power plant flue gas, and a detailed comparative cost analysis of this base case design is described in Sections 5 and 6. There are other ways to utilize sweep membrane units in CO_2 capture schemes, and Figure 29 shows two variations that use sweep membrane units in hybrid schemes with other non-membrane capture technologies. Figure 29(a) simply replaces the initial membrane unit proposed in Figure 25 with any other CO_2 capture technology (such as absorption or adsorption), using the sweep membrane unit in series with the selected capture technology. Another option is the Figure 29(b) process design, in which the sweep membrane unit is used in parallel with the CO_2 capture step.

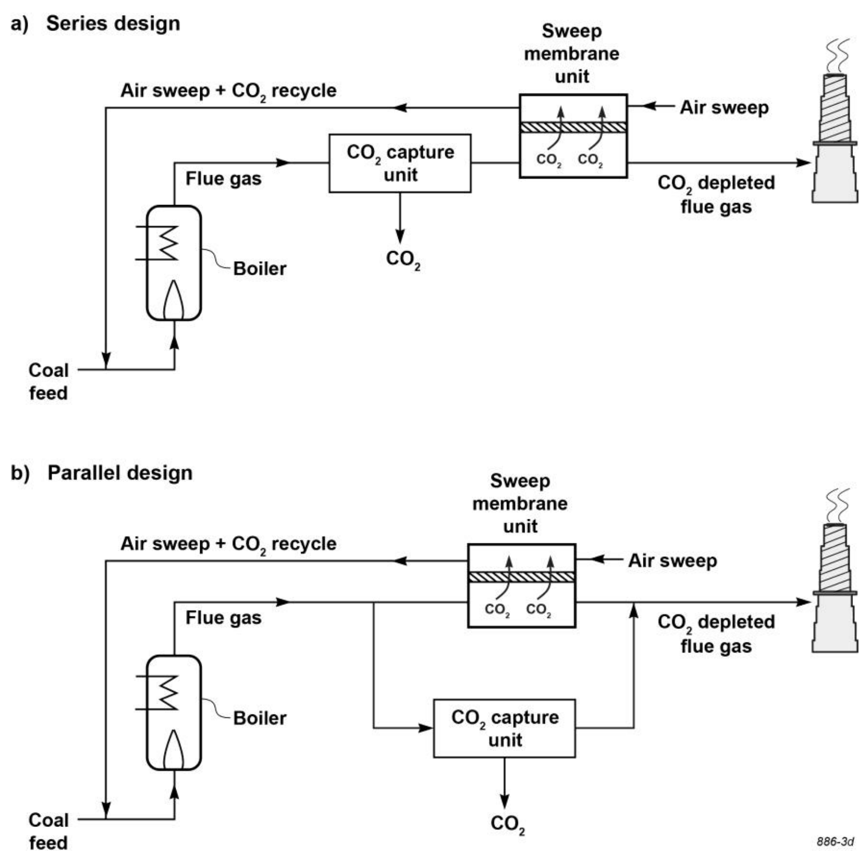


Figure 29. Simplified diagrams of proposed hybrid processes using membrane-based selective CO_2 recycle (a) in series and (b) in parallel with any other CO_2 capture technology, as applied to treatment of combustion flue gas.

A potential advantage of these hybrid designs is they do not require large compressors or vacuum equipment to capture CO_2 as the all-membrane process does. In the Figure 29(b) parallel design, the sweep membrane unit selective recycles CO_2, building up the concentration of CO_2 in the flue gas. A relatively small, concentrated CO_2 stream can then be sent to the capture unit operating in parallel with the sweep membrane. The size, cost, and energy use of the capture unit is greatly reduced compared to the case where it operates alone without the sweep membrane unit. This parallel hybrid approach looks particularly promising for CO_2 capture from natural gas-fired power plants and is described in more detail below.

Power generation by the natural gas combined cycle (NGCC) process generates less CO_2 than a coal-based process, due to the higher process efficiency and lower fuel carbon intensity. However, in the NGCC process, the air used is approximately 200-250% in excess of the stoichiometric ratio of air to fuel for complete combustion of natural gas. As a result, the CO_2 concentration in the flue gas is only 4%, or even less. Using the same methodology as used in Figures 27 and 28, we calculated the minimum energy required for CO_2 separation from NGCC flue gas, and compared the results with CO_2 capture from coal-fired flue gas, as shown in Figure 30. The difference in the energy required for CO_2 separation from these two processes is mainly caused by the different CO_2 concentrations in the flue gases. Due to the lower CO_2 concentration in the NGCC flue gas, it requires more energy to remove the CO_2 from this flue gas.

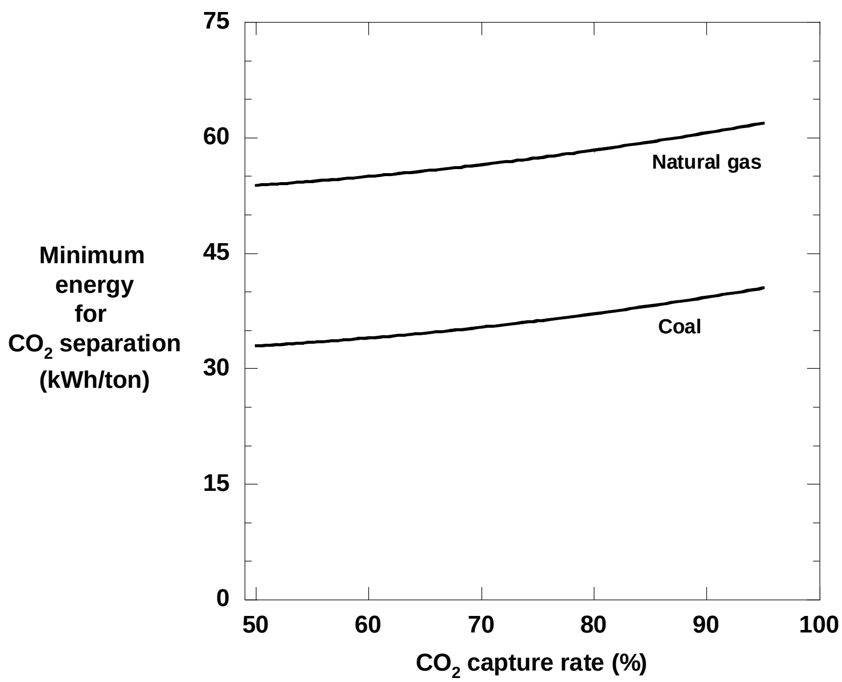


Figure 30. Comparison of minimum energy for CO_2 separation from natural gas and coal-fired power plant flue gases.

Figure 31 shows a parallel hybrid design for CO_2 capture from NGCC flue gas. Part of the flue gas from power generation is sent to a CO_2 capture unit. The rest of the gas enters the feed side of a countercurrent /sweep membrane module unit, and the CO_2-enriched permeate is selectively recirculated to the combustor and gas turbine. Because of the membrane selective CO_2 recycle, the CO_2 concentration in the flue gas increases significantly, and the flue gas stream fed to the capture unit is reduced in size, resulting in cost and energy savings for CO_2 capture. The amount that the CO_2 in the flue gas can be concentrated by selective CO_2 recycle is more dramatic for natural gas than for coal. For example, selective CO_2 recycle can approximately double the CO_2 concentration in coal flue gas (from 12% to 24%), but can increase NGCC flue gas CO_2 concentration by five times (4% to 20%). This difference is related to the much higher excess air used with natural gas (250%) compared to coal (15%). The high excess air used in a natural gas combustion turbine allows more flexibility to recycle additional CO_2 with a sweep membrane without starving the combustion process of oxygen. This additional CO_2 recycle permits greater CO_2 enrichment by sweep membrane for an NGCC process.

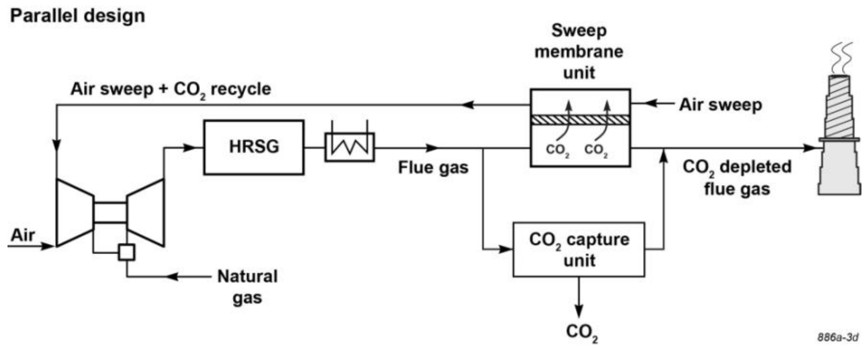


Figure 31. Simplified diagram of membrane-selective CO_2 recycle in parallel with a CO_2 capture step, as applied to CO_2 capture from NGCC flue gas.

One concern about the process described in Figure 31 is that the oxygen in the combustion air is diluted because of the recycle stream, which may affect the combustion conditions, and therefore, the emission and efficiency of the gas turbine. Figure 32 shows the correlation between CO_2 in the flue gas and oxygen in the combustion air when selective CO_2 recycle is used. In the calculation, the total mass of the gas entering the gas turbine was maintained at the same level as in the DOE base case study.[14] In another recent study conducted by GE researchers,[15] a bench-scale combustor integrated with a gas turbine was tested with oxygen-diluted combustion air by partially recycling the flue gas to the combustor. During the test, the oxygen concentration was fixed at 18.5%, and a stable flame was obtained. The GE researchers believe that, with minor modifications in the combustor design and its operation, it is feasible to run the combustor with 16% oxygen in the combustion air. As shown in Figure 32, when there is 16% oxygen in the combustion mixture, the selective CO_2 recycle approach can boost the CO_2 concentration in the flue gas to above 20%, a five-fold increase compared to the concentration with no sweep. For comparison, at the same oxygen concentration, the non-selective exhaust gas recycle (EGR) studied by GE and others produces about 7% CO_2 in the flue gas.

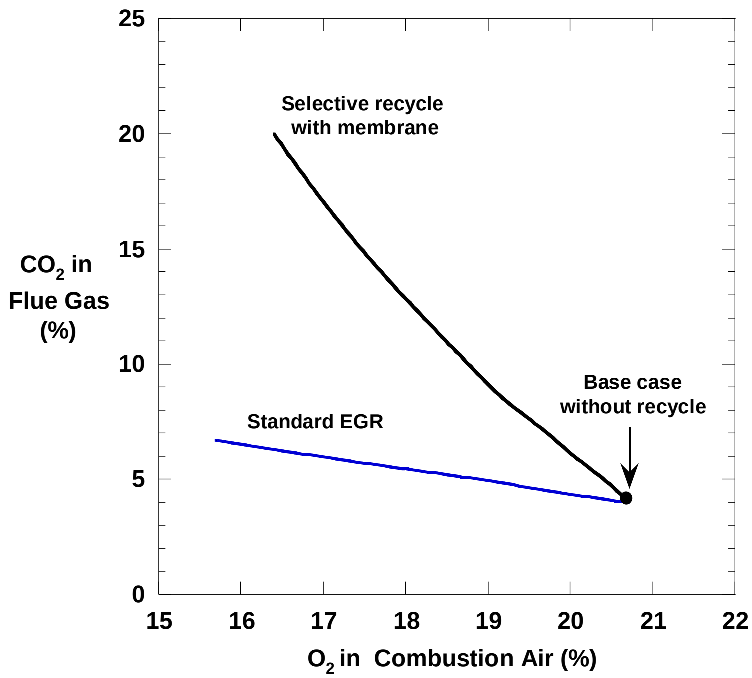


Figure 32. Correlation between CO_2 concentration in flue gas and O_2 concentration in combustion air when membrane-based selective CO_2 recycle is used in a NGCC process. For comparison, the correlation for standard non-selective exhaust gas recycle (EGR) is also shown.

Figure 33 shows the effect of selective CO_2 recycle on the minimum energy required for CO_2 separation from NGCC flue gas. The selective recycle can potentially reduce the minimum energy needed by 50%, due to CO_2 enrichment in the flue gas, when the oxygen in the combustion air is maintained at a level of 16%.

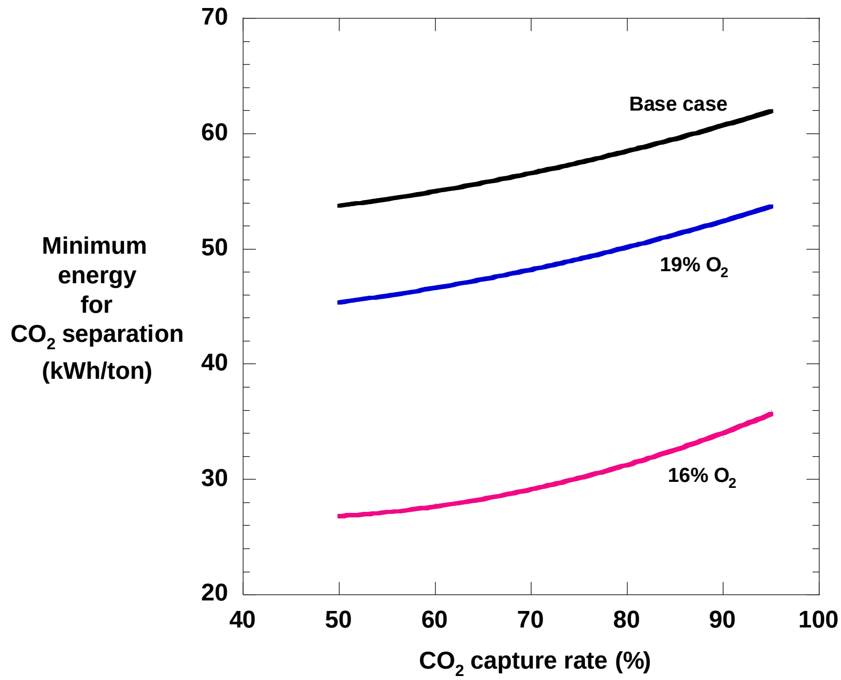


Figure 33. Effect of selective CO_2 recycle on the minimum energy required for CO_2 separation from NGCC flue gas.

As a final point of discussion on alternative parallel process designs, we have prepared an example that compares the use of amine absorption as the CO_2 capture process at power plants, with and without CO_2 recycle. The comparison shows the potential cost savings for CO_2 capture from NGCC flue gas that can be achieved by incorporating selective CO_2 recycle.

Table 3 compares the DOE NETL base case stream size and CO_2 concentration of the flue gas stream for coal-fired and NGCC power plants, and the energy and cost required for CO_2 separation from these streams by amine absorption.[14] Because the NGCC flue gas stream size is larger and CO_2 concentration is lower than for a coal-fired power plant, the cost and energy required for CO_2 separation is higher for NGCC flue gas in the base cases. The last column of Table 1 shows the comparable flue gas stream size and CO_2 concentration of the NGCC flue gas stream resulting from use of selective CO_2 recycle, as proposed in the Figure 31 parallel process design. Compared to the base case examples, the CO_2-enriched NGCC flue gas produced with selective recycle has a CO_2 concentration five times that of the NGCC base case and almost twice that of the coal base case. At the same time, the size of the stream to be treated is only about 20% that of the NGCC case, or 30% that of the coal-fired case. Therefore, the cost for CO_2 separation by amine absorption for the CO_2-enriched NGCC flue gas will be at most comparable to, and most likely less than, the cost for base case coal-fired flue gas when selective recycle is used.

Table 3. Comparison of Flue Gas Conditions and CO_2 Separation Costs for CO_2 Capture Using Amine Absorption at 550 MW_e Power Plants. Base case data are from DOE NETL[14]; selective recycle data are from simulations of the process shown in Figure 31.

Items for Comparison	Base Case Coal[a]	Base Case NGCC[a]	NGCC with Selective Recycle
CO_2 concentration in flue gas (%)	13.5%	4%	22.5%
Size of flue gas stream to the CO_2 capture unit (ton/hr)	2,138	3,231	618
Energy for CO_2 capture by amine absorption (kwh/ton)	400	440	<400
Cost for CO_2 separation by amine absorption ($/ton)	29[b]	43[b]	<29

a. Base case data taken from 2010 DOE NETL report [14]
b. Chapel, Mariz, Ernest, 1999 [16]

One additional factor that may increase the advantages to be realized from CO_2 selective recycle is the choice of other processes for the CO_2 capture unit. Amine absorption is a chemical-based absorption process. We believe that selective CO_2 recycle can provide even greater benefits to physical absorption/adsorption separation processes, where the process economics are mainly determined by the CO_2 partial pressure in the feed.

More quantitative assessments of the potential cost savings of parallel designs incorporating CO_2 selective recycle will be prepared in follow-on work.

5. BASE CASE DESIGN PERFORMANCE AND COST ANALYSIS

Introduction

One of the key objectives of this project was to conduct a performance and cost analysis of our membrane CO_2 capture process applied to an existing coal-fired power plant, and to compare these results with previous findings for conventional amine CO_2 capture technology. To conduct this analysis, MTR worked with EPRI and WorleyParsons (WP). WP focused on the technical aspects of the membrane retrofit and system integration at a power plant, as well as development of capital and operating cost estimates. EPRI used these costs to develop an economic analysis and comparison of CO_2 capture by the membrane process to the base case amine retrofit. MTR extended the EPRI analysis to include a sensitivity study that evaluates which future membrane performance or process improvements would be most beneficial to improve the competitiveness of membrane-based CO_2 capture.

The design and cost evaluation was based on retrofitting a membrane system for 90% CO_2 capture to American Electric Power (AEP) Conesville Unit No. 5. This plant was chosen for the design basis because amine-based CO_2 capture at this plant was documented in the DOE/NETL 401/110907 Report, entitled *Carbon Dioxide Capture from Existing Coal-Fired Power Plants*, November 2007.[17]

The membrane design chosen by MTR, EPRI and WP for this analysis was the base-case process shown in Figure 6 (Section 1) of this report, in which a membrane CO_2 capture step operates in series with a sweep membrane step. A pressure ratio of 10 was chosen for the membrane CO_2 capture step, corresponding to a feed pressure of 2 bar and a permeate vacuum of 0.2 bar. The vacuum pressure was selected based on a survey of equipment suppliers to identify the minimum pressure technically and economically feasible for an application of this scale. The feed pressure was chosen based on prior MTR design studies that showed a pressure ratio of 10 was a good balance between compression costs (which increase with increasing pressure ratio) and membrane costs (which decrease with increasing pressure ratio). The membrane properties used in the analysis are CO_2 permeance = 2,500 gpu and CO_2/N_2 = 50. We believe these values are achievable in the near term (1-3 years). The impacts of changing membrane properties or process design conditions, such as the pressure ratio, are described in the sensitivity analysis (Section 6) that follows the discussion of the EPRI/WP base case study.

EPRI/WP Base Case Evaluation Background

This section contains a summary of the background, methodology, and key findings of the EPRI/WP evaluation of the MTR base case design. The complete EPRI/WP findings are included in this report as the EPRI/WP Appendix.

Table 4 summarizes the cases that were compared in the EPRI/WP study. The cases highlighted with a light green background (SOA MEA Retrofit and Advanced MEA Retrofit) were developed in a standalone study described in more detail below.

It should be noted that there are many difficulties associated with comparing complex system costs from different studies that depend on many assumptions. A detailed discussion of these challenges is given in the EPRI/WP Final Report, Section 6.1. In light of these concerns, care should be taken in comparing the amine reference study to the membrane study, recognizing that each study has a cost estimate uncertainty of ±30%.

Table 4. Evaluation Matrix

Case	Description	CO_2 Capture/ Compression	Cost	Notes
Base-0	Do Nothing Case (Existing Facility)	None	NA	Existing Conesville Unit No. 5
MTR-1	MTR CO_2 Membrane Retrofit	90% capture/ 2015 psia	Dec 2009 $	Retrofit of Conesville Unit No. 5 [Focus of this Evaluation.]
MEA-1	MEA Retrofit Retrofit (SOA 2006)	90% capture/ 2015 psia	Escalate to Dec 2009 $	Solvent regeneration energy of 1550 Btu/lbm-CO_2.[17]
MEA-1a	MEA Retrofit Retrofit (Advanced)	90% capture/ 2015 psia	Cost presumed to be equivalent to MEA-1	Solvent regeneration energy of 1200 Btu/lbm-CO_2.[17]

a The MEA-1 and -1a Retrofit cases are known as “Case 1” and “Case 1a” within the DOE/NETL 401/110907 Report entitled *Carbon Dioxide Capture from Existing Coal-Fired Power Plants*, November 2007.[17]

Comparison Amine Study

A recent NETL study looked at the impact of retrofitting an existing PC-fired power plant with an amine-based CO_2 scrubbing system.[17] This study evaluated the technical, cost and economic impacts of removing CO_2 from the post-combustion flue gas of the Conesville Unit 5 Plant, using an advanced amine scrubbing system. The study evaluated various levels of CO_2 capture (0%, 30%, 50%, 70% and 90% capture), as well as providing a sensitivity study showing the effect of possible reductions in the solvent regeneration energy (1,550 and 1,200 Btu/lb-CO_2) for the 90% capture case. The 1,550 Btu/lb-CO_2 case represents the state-of-the-art MEA technology at the time of the study (~2006). The 1,200 Btu/lb-CO_2 level represents a near future value that may be achievable with improved solvent and other unspecified technological improvements.

The cost estimate for 90% CO_2 capture with amine-based scrubbing developed in the NETL study was escalated from July 2006 to December 2009 to enable comparison with the membrane capture system on an even basis. The escalation factors used were developed in-house at WP using various sources including industry publications, vendor inputs, and cost indices. Additional discussion regarding the MEA evaluation is presented in the EPRI/WP Final Report, Section 5.5.

Conesville Unit 5

Conesville Unit 5 is a nominal 450 MW_e reheat, subcritical, pulverized-coal (PC)-fired steam plant operated by AEP of Columbus, Ohio. Unit 5 is one of six coal-fired PC steam plants located on the Conesville site, with a total generating capacity of approximately 2,080 MW_e. The Unit 5 steam generator is a reheat unit with controlled circulation, a single furnace cell employing corner firing and tilting tangential burners. The fuel utilized is a bituminous coal from Ohio. The flue gas leaving the steam generator is cleaned by an electrostatic precipitator (ESP) and a lime-based flue gas desulfurization (FGD) system before being discharged to the atmosphere.

The steam turbine generator employs nominal steam conditions of 2,400 psig/1,000°F, exhausts to a condenser back pressure of approximately 2.5 inches Hga, and has been designed for 105% overpressure operation. The unit heat rejection is accomplished by two five-cell mechanical draft evaporative cooling towers.

The Conesville Unit 5 is representative in many ways of a large number of pulverized coal-fired units in use today throughout the United States. As such, it is an excellent unit for the subject of this CO_2 membrane retrofit study.

Modeling Approach

A critical input for determining the impact of the CO_2 membrane retrofit on Conesville Unit 5 is the development of the heat and mass balance and corresponding performance estimate. To this end, several different specialized computer modeling software programs were employed, each with its own niche in the overall analysis. The modeling software is listed below, followed by a brief description of how it was utilized within the analysis.

- *ChemCad using MTR proprietary membrane software* – MTR provided the performance data for each of the membrane units utilized in the base case process design. The software accounts for the membrane operating conditions, gas permeances, inlet composition, pressure ratio, sweep air flow rate, and geometry. The information provided by MTR was utilized by WorleyParsons in the supplemental analyses.
- *ASPEN* – WorleyParsons utilized ASPEN software to evaluate the impact of the membrane retrofit on the boiler, air and flue gas gaseous streams. The ASPEN analysis is complicated by the presence of two recycle streams: the sweep air from the counter-flow module, and the CO_2 purification system recycle stream. The presence of the recycle streams required that WorleyParsons and MTR iterate between their software models to ensure sufficient convergence of the results. Since the majority of the membrane retrofit impact affects the Conesville Unit 5 gas operation, the ASPEN analysis represents the heart of the overall WP analysis.
- *Boiler Performance Model (BPM)* – WorleyParsons utilized an internal BPM model to address the effect of the increased CO_2 and nitrogen flowing through the boiler as a result of the membrane's sweep-vitiated air feeding the secondary air fans, in lieu of fresh air. This simulation was not originally envisioned in the project scope, but was performed after the OEM for the steam generator declined to participate in the study. The BPM software was utilized primarily to determine an approximate impact of the membrane process on the boiler efficiency; it also provided a preliminary understanding of whether the existing temperature control schemes are adequate, in view of the redistribution of heat absorption within the boiler that results from the membrane retrofit.
- *GATE* – WorleyParsons utilized the GATE software to address the impacts to the steam turbine cycle resulting from integration of the membrane system. Initially, the GATE program was utilized to evaluate the process heat integration concepts, which were ultimately rejected as unjustified. By the end of the assessment, the only change to the steam turbine cycle was to account for a new steam extraction required by the CO_2 drying process.

The above models were used iteratively over the course of the study as the process assumptions were refined, and as the process models interfaced and converged. In addition, preliminary analyses were developed that helped address issues related to optimization of the overall process.

Membrane System Description

The retrofit of Conesville Unit 5 with the MTR membrane system for CO_2 capture requires the addition of new systems and modifications in the operation of the existing systems. A block flow diagram (BFD) of the retrofitted membrane system for CO_2 capture at Conesville Unit 5 is presented in Figure 34 on page 66; it forms the basis for the WP membrane CO_2 capture retrofit cost estimates. Twenty-seven new major items of equipment are identified by WP for the retrofit.[1] The following description of the retrofit process flow provides brief descriptions of the most important of these items.

Flue Gas Treatment Section. The first piece of newly installed equipment for the membrane retrofit is the heat exchanger ductwork between the ID fans and the FGD absorber (location Ⓐ for HX1-H from the major equipment list). Finned heat exchanger tubes are installed in the duct work to recover energy from the hot flue gas by transferring it into a circulating stream of glycol. Energy collected from the flue gas after it exits the ID fan is carried by the glycol to the exit of the counter-flow module B, where a second set of finned tubes heat the pressurized flue gas before it is fed to the flue gas expander. This transfer of energy allows the expander to achieve a greater power output, and will maintain additional thermal buoyancy in the flue gas exiting the stack. The lower flue gas temperature entering the FGD absorber also helps to reduce the amount of water that evaporates.

Flue gas, stripped of SO_2 in the FGD absorber, is fed to four compressors operating in parallel (location Ⓑ). These compressors raise the flue gas pressure up to the 2 bar design pressure of the MTR membranes. The pressurized flue gas is combined with residue from the MTR cross-flow module C (stream 43) and cooled in a direct contact cooler vessel (location Ⓒ). In the cooler, which is similar to an FGD absorber, cold water is sprayed over the gas to lower the temperature. Cold, pressurized flue gas is distributed by a header system to the banks of MTR cross-flow module A membranes (stream 22).

MTR CO_2 Membrane Section. MTR's membranes (locations Ⓓ and Ⓔ) capture CO_2 by using partial pressure as a driving force across a selective barrier material. The selectivity of the material allows a greater percentage of the CO_2 to preferentially permeate the membrane, while those compounds which would be impurities in a CO_2 product preferentially pass through as a residue stream. A vacuum on the permeate side of these cross-flow membranes (location Ⓕ) provides additional pressure gradient to drive the CO_2 capture.

The residue stream from the cross-flow module A membranes (stream 24) is distributed through banks of counter-flow module B membranes. In these counter-flow membranes (location E), the CO_2 permeates from the flue gas into the boiler's secondary air (stream 11). This membrane creates a CO_2 recirculation loop within the plant to ensure that the desired 90% CO_2 capture level is achieved. A booster air fan (location Ⓖ) will be installed to drive the secondary air through the banks of module B membranes. The CO_2-depleted flue gas which exits the module B membranes is at a pressure greater than that required to ensure proper dispersion through the plant stack (stream 27).

Flue Gas Treatment to Stack. The second set of finned tube heat exchanger tubes are installed in the ductwork between the cross-flow module B membranes and a flue gas expander (location Ⓗ for HX1-C from the major equipment list). These tubes transfer energy from the recirculating

[1] See Appendix C of the EPRI/WP Final Report for a full list of major equipment items. Appendix B of the same report provides more detailed flow diagrams for the new equipment sections.

glycol into the flue gas, raising the temperature of the gas. A single-stage expander (location Ⓘ) recovers energy from the hot pressurized gas as the pressure is reduced from the operating pressure of the membrane modules to the pressure required to dispel the gas through the stack.

Vacuum Pumps and CO_2 Compression System. Dry vacuum pumps are utilized to maintain vacuum on the permeate side of the cross-flow module A membranes (location Ⓕ). Liquid ring vacuum pumps, which are used at power plants to maintain the vacuum in the condenser, are not well suited to this membrane application. The CO_2 and SO_2 in the permeate stream are water-soluble gases, and water is used in high volumes as a sealant in a liquid ring vacuum pump. Dissolution of CO_2 into the water reduces the system capture percentage, and is undesirable in this application. In addition, auxiliary power consumption by liquid ring pumps is prohibitively expensive to the process. Alternatives to liquid ring vacuum pumps include dry-type compressors which are used to achieve large volumes of vacuum in the pulp and paper industry, and are better suited to the MTR membrane process.

Dry compressors operate more efficiently and consume less power. The vacuum pumps for this large-volume application would be designed with a combination of axial and radial flow stages to achieve the desired level of vacuum. Two large vacuum pumps are required by the process to capture permeate from MTR cross-flow module A. The CO_2-rich permeate from the MTR cross-flow module A also contains water and SO_2, which have an influence on the vacuum pump materials of construction. At the discharge of the vacuum pump system (stream 33), the CO_2-rich gas is cooled (location Ⓙ), water is removed (location Ⓚ), and the gas is piped to equipment that will perform further compression.

Three multi-stage compressors operating in parallel will be required to process the volume of gas exiting the vacuum pumps (location Ⓙ). Permeate from MTR's cross-flow module C (location Ⓜ) is introduced into one of the later stages of compression. Circulating water from a new cooling tower will be supplied to coolers after the compression stages, to remove heat which is generated by the compression process. Efficient intercooling reduces the auxiliary power consumption of the compressors. Pressurized CO_2-rich gas exiting the compressors must be dried and purified before being pumped to the final boundary limit pressure.

Dehydration and Purification Systems. A triethylene glycol (TEG) drying system is installed following the compression system to remove moisture which was not knocked out in the compression process (location Ⓚ). The TEG dryer is a temperature swing chemical absorption system in which lean and rich TEG solvent is circulated between the regenerator and absorber. For this application, in which the product gas requirement is less than 50 ppmv of water, a high dew point depression TEG dehydration process was selected.

Dry CO_2-rich gas still contains impurities which exceed the values specified by DOE and must be further treated. Purification of the CO_2-rich gas to produce CO_2 which meets the specified requirements is accomplished by utilizing a low-temperature partial condensation process integrated with a distillation column (location Ⓛ). Cooling water has already been utilized to reduce the temperature, and other heat sinks must be used. The gas exiting the CO_2 drying system (stream 36) is cooled down to the necessary temperature in two stages. The relatively hot

CO_2-rich gas leaving the drying system is used to meet the energy demands of the CO_2 stripping column reboiler in the first stage of the cooling, at the same time eliminating a process steam demand. A chiller system based on evaporation of liquid propane is used to reduce the temperature further and partially condense the CO_2-rich gas.

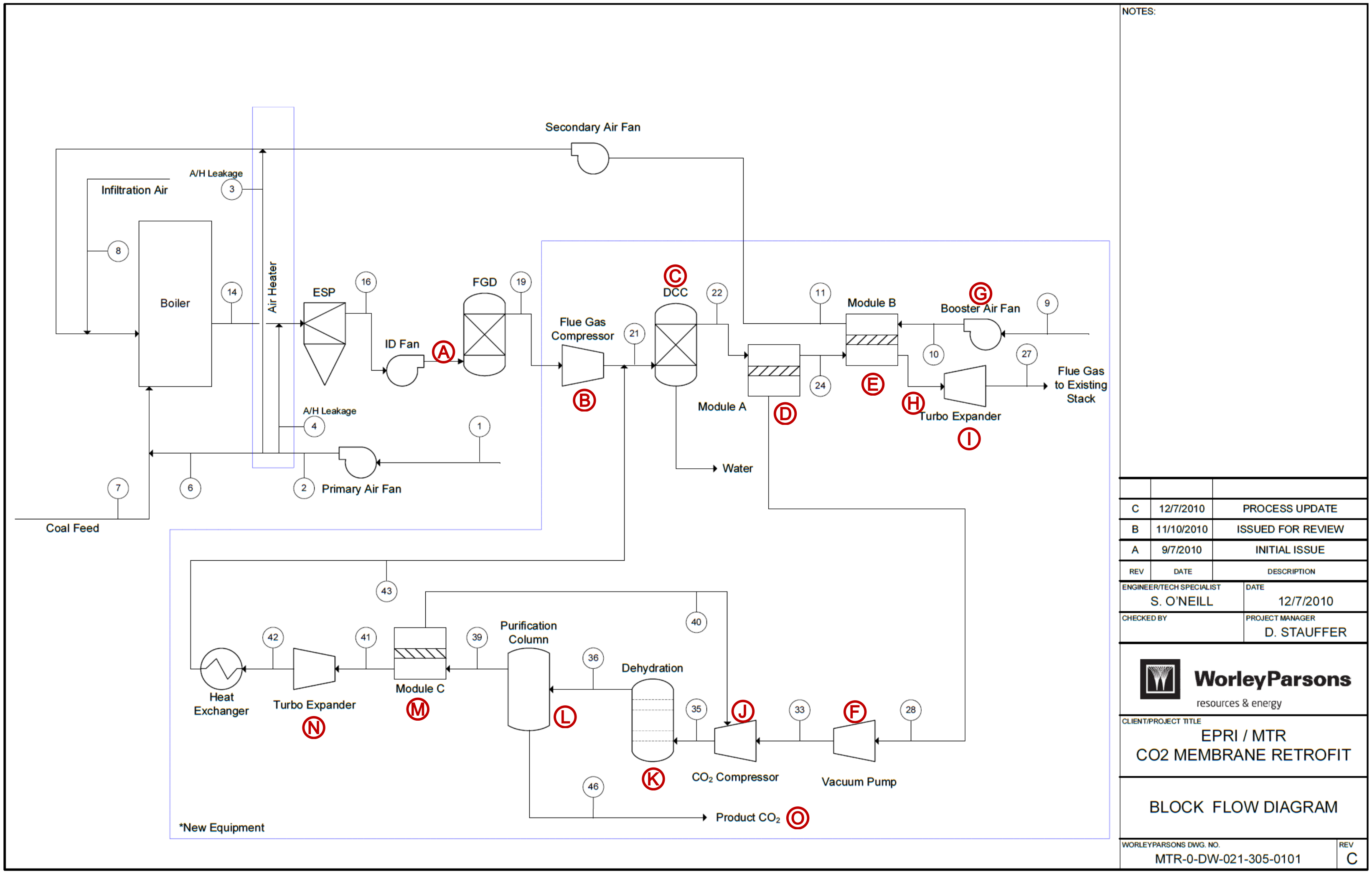


Figure 34. Block flow diagram of the retrofitted Unit 5.

Condensation of the CO_2 dominates at the design temperature and pressure of the gas condenser (HX3-H from major equipment list; part of equipment at location Ⓛ). Oxygen and nitrogen condense with the CO_2 in amounts that exceed the product specifications (see Appendix A, Section 2.4.2 for a discussion of CO_2 product specifications). Impurities in the CO_2 are removed by processing the liquid mixture in a stripping column. As the impure liquid CO_2 cascades down the column, vapor which is generated in the reboiler travels upwards. The impurities preferentially fractionate into the vapor phase as it moves up and out of the column. A pure liquid CO_2 product which meets all specifications is drawn off the bottom and pumped up to the final discharge pressure (stream 46).

Overheads from the CO_2 stripping column contain a residual fraction of CO_2 (stream 39). MTR cross-flow module C (location Ⓜ) recovers a portion of the CO_2 and returns it to the compression system (stream 40). The pressure differential between the column overheads and the suction pressure of the compressor stage is used as the driving force for permeation. The residue gas which passes through this cross-flow module (stream 41) is at high pressure and still contains CO_2. Power is recovered from the membrane retentate through a low-temperature expander (location Ⓝ). Expanded residue is reintroduced into the flue gas prior to the direct contact cooler (stream 43), to give the overall system another chance to capture the recycled CO_2.

Chilling System. A chilling system that utilizes propane as the refrigerant is used to achieve the temperatures necessary to condense the CO_2 mixture [see chiller package flow diagram in the EPRI/WP Final Report (Appendix B) for details]. Gaseous propane is compressed up to a pressure that will facilitate condensation at a temperature which can be achieved by cooling water. The efficiency of the chilling system is increased by using the liquid propane to reject heat to the purification process. Expanded module C retentate, CO_2 product, and stripping column overheads are all at temperatures lower than that of the liquid propane. Through heat exchange with those three gases, the liquid propane can be sub-cooled. This process reduces the losses associated with reducing the pressure of the liquid propane. At reduced pressure, the liquid-vapor mixture of propane is sent to the CO_2-rich gas condenser, where the liquid propane evaporates, inducing condensation in the CO_2-rich gas.

Cooling Water System. A new cooling tower and auxiliary cooling water system will be installed to meet the new process cooling demands (not shown). A set of circulating water pumps will be installed for the new tower. The circulating water will service the vacuum pumps, multi-stage compressors, direct contact cooler, and propane compressor systems. Make-up water demand for the new cooling tower will be offset by collecting the condensate from the DCC and the compression process and pumping it to the cooling tower basin.

Effect of Membrane Retrofit on Existing Plant Systems

The membrane retrofit uses secondary air as a sweep gas in counter-flow modules to enhance CO_2 capture. This results in the vitiated air being fed into the secondary air system, and also produces an increased mass and volumetric flow rate through the system, as described in Table 5.

Table 5. Changes in Secondary Air Due to Membrane Retrofit.

Parameter	Units	Base Case	Retrofit Case
Secondary Air Flow Rate			
Mass	lb/hr	2,843,126	3,539,302
Volumetric	MMscfd	901	1,071
CO_2 Content	Mol%	0.03	9.10
O_2 Content	Mol%	20.52	17.36
Average Molecular Weight	g/mol	28.74	30.11

These changes in secondary air flow rate and composition will potentially affect the performance of a number of power plant systems. WP performed a qualitative assessment of the impact of retrofitting a membrane system with sweep recycle on the existing power plant operation. Table 6 summarizes the results of this analysis.

Table 6. WP Summary of the Impact of Flue Gas Recycle on Performance of Power Plant Systems

System	WP Issues	WP Solution/Comments
Secondary Air System	Increased mass flow through the secondary air systems will increase the pressure drop through the ductwork.	A second forced draft (FD) fan must be used in parallel with the existing fan to meet the increased flow demand.
Flue Gas Handling System	The mass flow handled by the ID fans, as well as the amount of pressure head they must generate, is increased by the retrofit.	A second induced draft (ID) fan must be used in parallel with the existing fan to meet the increased flow demand.
Boiler Performance	• Extra CO_2 and N_2 introduced into the boiler with the secondary air sweep gas system are expected to reduce furnace temperature and steam generation, while flue gas velocity is expected to increase. Heat absorption in the boiler will be shifted to the boiler back pass and superheated steam temperature could increase. • The reduced O_2 content in the secondary air is likely to increase the unburned carbon (UBC) in the ash. The UBC affects the boiler efficiency and CO levels.	• An assumed solution to shift heat absorption back to the furnace is to tilt the burners down, correcting both steam generation and steam temperature. • The UBC may be mitigated by grinding of the pulverized coal to a finer size. This is analogous to the pulverizer modifications required for low NO_x burner retrofits. • The membrane retrofit is expected to result in a slightly greater boiler dry gas loss. Hence, efficiency of the retrofitted boiler is expected to be reduced by approximately 0.8%.
Combustion System	Since the O_2 content in the primary air and the primary air temperature will remain unchanged with the CO_2 membrane retrofit, no burner modifications are expected. However, the secondary air distribution between boiler windboxes, closed coupled over-fire air (CCOFA) and SOFA may need to be adjusted with increased secondary air flow rate and reduced O_2 content.	Assessment of the impact on windbox/CCOFA/SOFA operation requires detailed analysis and testing by the boiler vendor.
Boiler Back Pass	With an increased boiler flue gas flow rate, there is a significant risk of increased erosion of the boiler back pass surfaces.	Detailed analysis by a boiler OEM may recommend duct modifications to minimize erosion.
ESP	Increased volumetric flow will decrease the contact time between the flue gas and ESP and will reduce the collection efficiency of the ESP.	The newly added DCC and membrane system will more than make up for the reduced ESP efficiency, and the emitted PM will be lower than pre-retrofit.
FGD	Increased volumetric flow may lower FGD SO_2 removal performance.	Lower FGD performance will not be an issue because the membranes are permeable to both SO_2 and SO_3, and will further reduce the sulfur levels in the flue gas by over 90%.
Steam Cycle	Electrical energy is the prime driver to create pressure ratio for CO_2 capture with membranes, so there is minimal effect on the steam cycle.	None.
Electric System	New equipment associated with the membrane system – compressors, pumps, gas expander generators – will require a new medium voltage 13.8 kV system to support operation of motors larger than 5,000 HP.	The scope of modifications to the auxiliary power distribution system is envisioned to include the addition of segregated-phase bus ducts, switchgear and control equipment, service transformers, generator equipment, station service equipment, conduit and cable trays, wire, and cable with all required foundations, and standby equipment.

In summary, the membrane retrofit appears feasible from a plant operations standpoint, with appropriate system engineering. The biggest concern for the membrane retrofit is the impact on boiler performance of using secondary air as a sweep gas to recycle CO_2 to the boiler. It is estimated that this process will reduce boiler efficiency by 0.8%. It is recommended that a more detailed analysis be conducted by a boiler manufacturer to refine this estimate. If a value in this range is confirmed, MTR believes the benefits of sweep recycle to CO_2 capture will significantly outweigh the reduction in boiler efficiency.

Membrane System Layout

The availability of space at a power plant is a factor in the installation of a CO_2 capture system. This is true at Conesville where limited space is available in the vicinity of the unit being retrofitted. Because the membrane system would be negatively affected by gas pressure drops if the equipment was far removed from the existing boiler structures, a multi-story building to house the membranes is designed adjacent to the Unit 5 FGD absorber vessels.

Large rotating equipment is most effectively installed at ground level. Ground level installation allows for the construction of rigid foundations capable of supporting the weight and vibrations of the rotating equipment. Membranes are installed on a floor above the various compressors in order to minimize ductwork and pressure drop.

Captured dilute CO_2 gas is ducted to a separate area of the plant for processing. Sufficient open land is available to facilitate the outdoor installation of the CO_2 compression and purification systems. A new cooling tower capable of servicing all of the new equipment is co-located with the CO_2 compression and purification equipment. A proposed equipment layout drawing is provided in the EPRI/WP Final Report, Section 4.4.

Membrane System Performance Summary

With the installation of new CO_2 capture equipment and the change in the operating conditions of the existing plant, the net power generated by the plant is reduced. The retrofitted Unit 5 performance is summarized in Table 7.

Table 7. WP Evaluation of MTR Base Case Membrane System: Retrofitted Conesville Unit 5 Performance Summary.

Equipment	Units	Retrofit Case
Steam turbine generator	kW	463,044
Flue gas expander generator	kW	21,018
Recycle gas expander generator	kW	2,834
Gross power generation	kW	486,896
Existing plant auxiliary loads	kW	33,430
CO_2 capture process auxiliary loads	kW	142,924
Net power generation	kW	310,542

Individual auxiliary loads for the newly installed equipment are presented in Table 8.

Table 8. WP Evaluation of MTR Base Case Membrane System: Retrofitted Conesville Unit 5 Auxiliary Loads.

Equipment	Units	Retrofit Case
Booster air fan	kW	3,721
Flue gas compressors	kW	54,610
Vacuum pumps	kW	24,063
CO_2 compressors	kW	42,649
CO_2 dryer	kW	133
Chiller compressors	kW	13,009
CO_2 pump	kW	2,282
Auxiliary cooling service	kW	2,457
Total auxiliary load	kW	142,924

All of the equipment required for the efficient operation of the MTR CO_2 capture process is not currently commercially available. Turbomachinery vendors who were contacted for this study indicated that while there may not currently be off-the-shelf equipment for this particular application, the potential exists to engineer the required equipment from industry standard designs. Vendors also indicated a willingness to develop the equipment should the market develop.

As a result of the retrofitted CO_2 capture system, the composition of the flue gas that is discharged from the stack will change. WP assumed the temperature and pressure of the flue gas entering the stack was held constant to allow for the necessary buoyancy to carry the gas out of the stack and ensure sufficient mixing in the atmosphere. Changes in the composition of the flue gas are presented in Table 9.

Table 9. Stack Gas Composition

Constituent	Units	Base Case	Retrofit Case
Argon	mol %	0.80	1.08
Carbon dioxide	mol %	12.56	1.69
Water	mol %	16.36	1.24
Nitrogen	mol %	67.08	90.93
Nitrous oxides	mol %	0.02	0.03
Oxygen	mol %	3.18	5.03
Sulfur dioxide	mol %	0.00 (ca 50 ppm)	0.00 (ca 2 ppm)

In addition to capturing CO_2, the membranes also reduce the stack emissions of NO_x and SO_2 because these species permeate through the membrane. Post-retrofit reductions in NO_x and SO_2 emissions are illustrated in Table 10.

Table 10. NO_x and SO_x Emissions

Constituent	Units	Base Case	Retrofit Case
Nitrous oxides	tons/day	15.1	11.7
Sulfur dioxide	tons/day	6.3	0.2

The primary goal of the membranes is to capture CO_2. Sufficient membrane area is installed so that nominally 90% of the CO_2 generated is captured. Reduction in the rate of plant CO_2 emissions is presented in Table 11.

Table 11. CO_2 Emissions

Constituent	Units	Base Case	Retrofit Case
Carbon dioxide	lbs/hr	866,102	90,007
Carbon dioxide	tons/day	10,393	1,080

The product CO_2 purity achieved by the membrane and purification systems meets or exceeds the specified CO_2 quality, as presented in Table 12.

Table 12. Product CO_2 Composition

Parameter	Units	Specified Value	Retrofit Case
Pressure	psia	2,015	2,015
CO_2, min	vol%	96%	99.98%
H_2O, max	vol%	0.015%	0.000%
N_2, max	vol%	0.6%	0.01%
O_2, max	ppmv	100	100
SO_2, max	vol%	1%	0% (< 1 ppm)

Note: The composition values listed for the retrofit case are mol%.

Membrane System Cost Analysis

The basis for capital and operating cost estimates in this study is consistent with the basis in the 2007 DOE/NETL *Carbon Dioxide Capture from Existing Coal-Fired Plants* study, except that the cost analysis in this report is expressed in December 2009 U.S. dollars. This approach enables comparison of the results from this study with the appropriately escalated results of the 2007 DOE/NETL study.

The capital cost estimates in this study are assessed on a Total Investment Cost (TIC) level,[2] and are presented on an engineering, procurement, and construction (EPC) reimbursable basis, with

[2] The TIC cost documented in the reference study is consistent with the Total Plant Cost (TPC) nomenclature utilized by DOE/NETL.

process and project contingencies. All costs are estimated in December 2009 U.S. dollars. These costs include all required equipment to complete the retrofit, such as the new membrane-based CO_2 capture system, the new CO_2 compression and dehydration system, additional new balance of plant systems, and modifications to the existing plant equipment and systems as required to support operation of the retrofitted plant.

Operating and maintenance (O&M) costs were calculated for all systems. The O&M costs for the Base Case (pre-retrofit Conesville #5 Unit) are based on the 2007 DOE/NETL study. For the retrofit CO_2 capture system evaluations, additional O&M costs are calculated for the new equipment. The variable operating and maintenance (VO&M) costs for the new equipment include such categories as chemicals, membrane replacement, maintenance material and labor, and make-up power cost (MUPC) from the reduction in net electricity production. The fixed operating and maintenance (FO&M) costs for the new equipment include operating labor and maintenance.

A detailed discussion of the assumptions and methodology used in the cost analysis is provided in the EPRI/WP Final Report, Section 5.

Capital and O&M Cost Results

The capital cost summary for the MTR retrofit case is presented in Table 13. Additional cost details for this plant are available in the EPRI/WP Final Report, Section 5.3.

The total plant cost for the membrane retrofit at 90% CO_2 capture is $594 million. Of this cost, more than 50% is for compression equipment required to pressurize the flue gas feed to 2 bar and to compress the captured CO_2 to 2,015 psia. A detailed sensitivity study examining the impact of compression costs and membrane process design on total plant cost is provided in Section 6 of this report.

Table 13. Total Plant Cost for MTR-1: Cost Summary

Client:	DOE/NETL/BAH									**Report Date:**	2010-Dec-23	
Project:	MTR CO2 Membrane for Capturing CO2 from Power Plant Flue Gas											
	TOTAL PLANT COST SUMMARY											
Case:	MTR CO2 Membrane Retrofit											
Plant Size:	391.2 TPH CO2		**Estimate Type:**		Conceptual				**Cost Base (Dec)**	2009	($x1000)	
Acct No.	**Item/Description**	**Equipment Cost**	**Material Cost**	**Labor Direct**	**Labor Indirect**	**Sales Tax**	**Bare Erected Cost $**	**Eng'g CM H.O.& Fee**	**Contingencies Process**	**Contingencies Project**	**TOTAL PLANT COST $ x 1000**	**TOTAL PLANT COST $K / TPH**
1	COAL & SORBENT HANDLING	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
2	COAL & SORBENT PREP & FEED	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
3	FEEDWATER & MISC. BOP SYSTEMS	$3,657	$0	$1,910	$0	$0	**$5,567**	$516	$0	$1,217	**$7,300**	$19
4	PC BOILER & ACCESSORIES	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
5A	FLUE GAS TREATMENT	$64,173	$25,106	$51,271	$0	$0	**$140,550**	$13,542	$0	$30,473	**$184,565**	$472
5B	CO2 PURIFICATION & COMPRESSION	$122,359	$10,891	$57,103	$0	$0	**$190,353**	$17,893	$8,400	$43,329	**$259,974**	$665
6	COMBUSTION TURBINE/ACCESSORIES	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
7	HRSG, DUCTING & STACK	$6,378	$1,959	$10,121	$0	$0	**$18,457**	$1,647	$0	$3,211	**$23,315**	$60
8	STEAM TURBINE GENERATOR	$18	$0	$70	$0	$0	**$88**	$8	$0	$19	**$115**	$0
9	COOLING WATER SYSTEM	$6,508	$2,965	$6,492	$0	$0	**$15,966**	$1,480	$0	$2,462	**$19,908**	$51
10	ASH/SPENT SORBENT HANDLING SYS	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
11	ACCESSORY ELECTRIC PLANT	$9,747	$6,166	$26,205	$0	$0	**$42,118**	$3,896	$0	$6,080	**$52,094**	$133
12	INSTRUMENTATION & CONTROL	$5,887	$0	$9,153	$0	$0	**$15,040**	$1,405	$0	$2,041	**$18,487**	$47
13	IMPROVEMENTS TO SITE	$1,953	$1,123	$5,746	$0	$0	**$8,823**	$866	$0	$1,938	**$11,627**	$30
14	BUILDINGS & STRUCTURES	$6,364	$505	$6,758	$0	$0	**$13,627**	$1,242	$0	$2,230	**$17,100**	$44
	TOTAL COST	**$227,044**	**$48,715**	**$174,829**	**$0**	**$0**	**$450,588**	**$42,496**	**$8,400**	**$93,000**	**$594,484**	$1,520

The operating and maintenance cost estimates for the MTR-1 case are presented in Table 14. The total variable operating costs are $73 million/year; the bulk of this cost is supplemental electricity to run the CO_2 capture equipment.

Table 14. Operating & Maintenance Cost for MTR-1

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Dec):	2009
MTR CO2 Membrane Retrofit					Heat Rate-net (Btu/kWh):	13,762
					CO2 TPH:	391
					Capacity Factor (%):	85
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate (base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	0.0		0.0			
Operator	2.0		2.0			
Foreman	0.0		0.0			
Lab Tech's, etc.	0.0		0.0			
TOTAL-O.J.'s	2.0		2.0			
					Annual Cost $	Annual Unit Cost $/TPH
Annual Operating Labor Cost					$789,188	$2.018
Maintenance Labor Cost					$2,956,092	$7.557
Administrative & Support Labor					$936,320	$2.394
TOTAL FIXED OPERATING COSTS					**$4,681,601**	**$11.969**
VARIABLE OPERATING COSTS						
						$/TPH
Maintenance Material Cost					**$4,434,138**	**$0.00152**
Consumables	Consumption Initial	 /Day	Unit Cost	Initial Cost		
Water (/1000 gallons)	0	3,744.00	1.23	**$0**	**$1,428,423**	**$0.00049**
Chemicals						
MU & WT Chem.(lbs)	63,432	9,061.68	0.20	$12,481	$553,158	$0.00019
Replacement Membrane Modules (m2)	0	498.88	10.00	$0	$1,547,779	$0.00053
Subtotal Chemicals				**$12,481**	**$2,100,938**	**$0.00072**
Other						
Supplemental Fuel (MBtu)	0	0	0.00	$0	$0	$0.00000
Supplemental Electricity (for consumption) (MWh)	0	2,958	71.00	$0	$65,150,683	$0.02237
Subtotal Other				**$0**	**$65,150,683**	**$0.02237**
Waste Disposal						
Fly Ash (ton)	0	0.00	18.45	$0	$0	$0.00000
Bottom Ash (ton)	0	0.00	18.45	$0	$0	$0.00000
Subtotal-Waste Disposal				**$0**	**$0**	**$0.00000**
By-products & Emissions						
CO2 Product (ton)	0	9,388	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$12,481**	**$73,114,182**	**$0.02510**
Fuel (ton)	0	0	0.00	**$0**	**$0**	**$0.00000**

Membrane System Performance Comparison

Key technical performance parameters for the MTR CO_2 membrane retrofit are compared to the base case (status quo), and to the MEA-1 (SOA 2006) and MEA-1A (Advanced MEA Technology) cases in Table 15.

Table 15. Summary Comparing Technical Performance for Retrofitting Conesville Unit 5 Using Different Technologies.

Parameter	Units	Case-0	MTR-1	MEA-1	MEA-1A
		Base Case	SOA 2010	SOA 2006	Advanced
Boiler Parameters					
Main Steam Flow	lbm/hr	3,131,619	3,131,600	3,131,651	3,131,651
Reheat Steam Flow (to IP turbine)	lbm/hr	2,853,607	2,851,724	2,848,739	2,848,725
Main Steam Pressure	psia	2,535	2,535	2,535	2,535
Main Steam Temp	Deg F	1,000	1,000	1,000	1,000
Reheat Steam Temp	Deg F	1,000	1,000	1,000	1,000
Boiler Efficiency	percent	88.13	87.33	88.13	88.13
Flue Gas Flow leaving Economizer	lbm/hr	4,014,743	4,795,700	4,014,743	4,014,743
Flue Gas Temperature leaving Air Heater	Deg F	311	322	311	311
Coal Heat Input (HHV)	MMBtu/hr	4,228.7	4,273.6	4,228.7	4,228.7
Coal Heat Input (HHV)	lbm/hr	374,453	378,425	374,453	374,453
CO2 Removal System Steam & Related Parameters					
Solvent Regeneration Energy	Btu/lbm-CO2		**NA**	**1,550**	**1,200**
CO2 Removal System Steam Pressure	psia	---	203	47	47
CO2 Removal System Steam Temp	Deg F	---	718	424	424
CO2 Removal System Steam Extraction Flow	lbm/hr	---	5,696	1,210,043	975,152
CO2 Removal System Heat to Cooling Tower	MMBtu/hr	-	1,075	890.2	698.2
Natural Gas Heat Input	MMBtu/hr	-	-	13	13
CO2 produced from Natural Gas usage	lbm/hr	-	-	1,492	1,492
Generation & Auxiliary Load					
Existing Steam Turbine Generator Output	kW	463,478	463,044	342,693	367,859
CO2 Removal System Turbine Generator Output	kW	-	23,852	45,321	36,083
Total Turbine Generator Output	kW	463,478	486,896	388,014	403,942
Auxilliary Power: Existing Plant	kW	29,700	33,430	29,765	29,817
Auxilliary Power: CO2 Removal System	kW	-	142,924	54,939	54,845
Net Plant Output	**kW**	**433,778**	**310,542**	**303,310**	**319,280**
Plant Performance Parameters					
Net Plant Heat Rate (HHV)	Btu/kWh	9,749	13,762	13,985	13,285
Net Plant Efficiency (HHV)	%	35.01%	24.80%	24.41%	25.69%
Energy Penalty, (percentage points of NP Eff.)	%	Base	10.21%	10.60%	9.32%
Capacity Factor	%	85%	85%	85%	85%
Plant CO2 Emissions					
Carbon Dioxide Produced	lbm/hr	866,102	872,189	867,595	867,595
Carbon Dioxide Recovered	lbm/hr	-	782,177	779,775	779,775
Carbon Dioxide Emissions	lbm/hr	866,102	90,012	87,820	87,820
Carbon Dioxide Recovered (% of Produced)	%	0.00%	89.68%	89.88%	89.88%
Specific Carbon Dioxide Emissions	lbm/kWh	1.997	0.290	0.290	0.275
Normalized Specific CO2 Emissions (Relative to Base Case)	fraction	1.000	0.145	0.145	0.138
Avoided Carbon Dioxide Emissions (as compared to Base)	lbm/kWh	---	1.707	1.707	1.722

Note: The source of values for Base Case, MEA-1, and MEA-1A can be found within the DOE/NETL 401/110907 Report entitled *Carbon Dioxide Capture from Existing Coal-Fired Power Plants*, November 2007.[17]

The boiler parameters section shows that the main steam (MS) and reheat steam (RH) flow rates are nearly identical in all cases. The MS and RH temperatures are controlled to 1,000°F in all cases. The boiler efficiency of the MTR-1 case is 0.8% lower than the other comparison cases.

The boiler efficiency of the MEA cases are unaffected compared to the base case, while the MTR-1 case has a lower efficiency as a result of the increased N_2 and CO_2 traveling through the boiler, yielding lower temperatures in the furnace area and a higher flue gas temperature exiting the air heater. The decreased boiler efficiency yields a coal flow increase of about 1%.

A comparison of the CO_2 removal systems steam parameters shows that the MEA cases use a very large amount (1.0 to 1.2 million lb/h) of low pressure steam for solvent regeneration, compared to a small amount (about 6,000 lb/h) of low pressure steam for the MTR case for CO_2 product drying. The MEA cases utilized natural gas for CO_2 drying. The heat rejection for the MTR case is higher than the MEA cases; it represents heat not only from the CO_2 compressors but also heat from the purification system chiller.

A comparison of the generation and auxiliary loads reveals that the MTR-1 steam turbine generator is nearly unchanged from the Base Case. The small change is a result of the steam extraction required for CO_2 drying. In contrast, the MEA cases are nearly 100 to 120 MW lower in gross generation because of the steam extraction for solvent regeneration. Both the MTR-1 and the MEA cases have additional generation related to the CO_2 removal system. The total gross generation for the MTR-1 case is nearly 100 MW higher than the MEA-1 case. The auxiliary load of the existing plant for the MTR-1 case increases slightly, from approximately 29,700 kW to 33,430 kW. This results primarily from the increased fan load on the secondary air fan due to both a 20% higher flow and a notable increase in pressure head. The largest change in this section is the auxiliary load for the CO_2 removal system, which for the MTR-1 case is nearly 143 MW, while the MEA-1 and -1A cases are nearly 55 MW. What the MTR case gained in increased gross generation, it lost back in an increased auxiliary load. Said differently, instead of having a large steam auxiliary load such as in the MEA case, the MTR case has a large electric auxiliary load.

The MTR-1 Case yields a net generation of 310,542 kW, which is higher than the MEA-1 case value of 303,310 kW, but less than the Advanced MEA-1A case value of 319,280 kW. Advanced membrane cases that produce higher net generation are discussed in the following section (Section 6).

Because the coal input only varies by a single percent between all the cases, the net plant efficiency trend simply echoes that of the net plant generation. The MTR-1 case produced about 0.7% more CO_2 than the other cases because of the decreased boiler efficiency. All capture cases capture 90% of the CO_2 produced in the boiler.

Incremental capital and O&M costs for the MTR CO_2 membrane retrofit are compared to the base case (status quo), and to the MEA-1 (SOA 2006) case in Table 12. The MTR-1 case shows a higher capital cost than the MEA-1 case ($594 million versus $475 million), but lower O&M costs ($77.8 million/yr for MTR-1 compared to $92.3 million/yr for MEA-1).

It should be noted that there are many difficulties associated with comparing complex system costs from different studies that depend on many assumptions. A detailed discussion of these challenges is given in the EPRI/WP Final Report, Section 6.1. In light of these concerns, care should be taken in comparing the amine reference study to the membrane study. Likewise, care should be taken in interpreting a 20-25% cost difference as an absolute difference, when each study has a cost estimate uncertainty of ±30%.

Table 16. Incremental Capital and O&M Cost Comparison for Retrofitting Conesville Unit 5.

Parameter	Units	Case-0	MTR-1	MEA-1	MEA-1
			Dec 2009 $	Jul 2006 $	Dec 2009 $
Capital Costs					
Bare Erected Cost	$1,000	Base	450,588	217,189	258,382
Eng, CM, HO, Fees, etc.	$1,000	Base	42,496	87,789	103,772
Project Contingency	$1,000	Base	8,400	56,022	66,429
Process Contingency	$1,000	Base	93,000	39,094	46,357
Total Investment Cost	**$1,000**	**Base**	**594,484**	**400,094**	**474,940**
Total Investment Cost	$/kW,net	Base	1,914	1,319	1,566
Operating & Maintenance Costs					
Fixed O&M Costs	$1,000/yr	Base	4,681	2,494	2,647
Variable O&M Costs	$1,000/yr	Base	7,963	17,645	20,631
Levelized, Makeup Power Cost	$1,000/yr	Base	65,151	62,194	68,996
CO2 Byproduct Revenue	$1,000/yr	Base	-	-	-
Feedstock (Natural Gas) O&M Cost	$1,000/yr	Base	-	653	575

Note: Costs for MEA-1 (Y2006 USD) are based on information in the DOE/NETL 401/110907 Report entitled *Carbon Dioxide Capture from Existing Coal-Fired Power Plants*, November 2007.[17]

Using the capital and O&M costs described above, EPRI calculated incremental LCOE values for the base case MTR membrane system retrofitted for 90% CO_2 capture on Conesville Unit 5. The methodology used for these calculations is described in the EPRI LCOE Analysis of the EPRI/WP Appendix. Figure 35 shows the incremental LCOE values for the MTR base case (MTR) and several variations of this case where equipment costs or system performance were changed. For the base case membrane system, the incremental LCOE is $57.1/MWh. Changes to compressor costs (±20%), membrane equipment costs (±20%), CO_2 product purity, and compression efficiencies all produce relatively minor variations in the incremental LCOE (±10%). The greatest impact on LCOE is achieved by using advanced membranes that are permeable enough to operate without feed compression. For this case, the capital cost is greatly reduced because feed compressors and residue turbo-expanders are no longer needed. The resulting incremental LCOE is reduced to $44/MWh. This compares to $55.6/MWh for the MEA-1 process and $53/MWh for the Advanced MEA-1A case. Further details on the advanced membrane case are provided in Section 6.

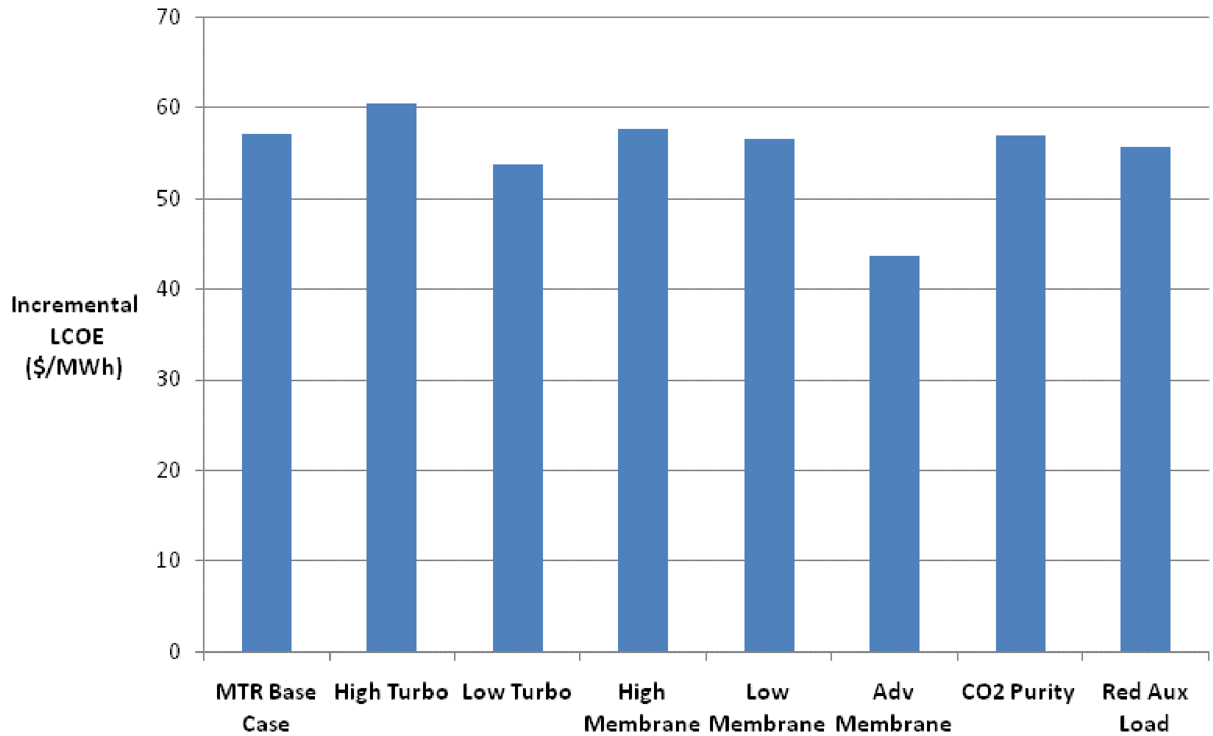


Figure 35. Incremental LCOE for 90% CO_2 capture from Conesville Unit 5 using variations of the MTR base case capture system. High and Low Turbo refer to cases with ±20% in compressor costs; High and Low Membrane are cases with ±20% in membrane costs; Adv Membrane is a case where next-generation membranes with no feed compression are used; CO_2 Purity is a case where the oxygen limit of 50 ppm in the CO_2 product is relaxed; Red Aux Load refers to a case where high compression efficiencies (93%) are used.

EPRI/WP Base Case Study Summary

The systems and economic analysis conducted by EPRI/WP confirmed that the MTR membrane process could be retrofitted to a coal-fired power plant and achieve 90% CO_2 capture. The estimated capital cost of this membrane system is 25% more than that of the base case MEA-1 system, while the energy use and O&M costs are slightly lower (~15%) for the membrane process. The resulting incremental LCOE for the base case membrane system is approximately equivalent to that of the MEA-1 process. With the ±30% uncertainty of the cost estimates in this study, it can be concluded that this base case membrane process is competitive with a state-of-the-art amine CO_2 capture system.

The primary reason for the relatively high capital cost of the membrane system is the large cost associated with the compression equipment required to generate a driving force for membrane CO_2 permeation. The impact of lower compression costs, as well as alternative membrane process designs that mitigate compression requirements, are evaluated in Section 6 of this report.

6. SENSITIVITY ANALYSIS AND REVIEW OF OTHER COMPETITIVE ISSUES

Introduction

The previous section of this report described a detailed system and cost analysis of a base case membrane process being retrofitted to the Conesville Unit 5 boiler to accomplish 90% CO_2 capture from the process flue gas. The analysis, prepared by EPRI/WP, concludes that membranes can be technically competitive with amines in CO_2 treatment at power plants, and show potential for lower energy use and operating costs. However, at the current stage of system development, the analysis shows that the capital cost of the base case membrane system may be higher than that of a state-of-the-art (SOA) amine system. This section provides a sensitivity analysis of the retrofit design, to understand the key factors that affect the cost and performance of a membrane CO_2 capture system. This analysis helps clarify which future membrane performance, process improvements, or other factors would be most beneficial to improving the cost competitiveness of membrane-based CO_2 capture. The latter portion of this section briefly reviews other factors (capture process chemicals disposal and emissions, system footprint, and so forth) that may affect the choice of CO_2 capture technology, but that were not quantified in this report.

The Impact of Compression Costs

Figure 36 shows the breakdown of EPRI/WP estimated capital costs for the base case membrane system designed for 90% CO_2 capture from Conesville Unit 5. In this estimate, more than 50% of the membrane system cost is compression equipment (~$300 million). This equipment includes a feed gas compressor, a residue stream turbo-expander, an air sweep blower, a permeate vacuum pump, and CO_2 compressors. After the compression equipment, the most costly items are the electric plant required to run the compressors, the CO_2 liquefaction chiller, buildings and instrumentation, and the membrane skids – each amounting to slightly less than 10% of the total system cost. This cost estimate suggests that the competitiveness of the base case membrane system hinges on compression costs. For this reason, we reviewed the compression cost assumptions to see what reductions might be possible.

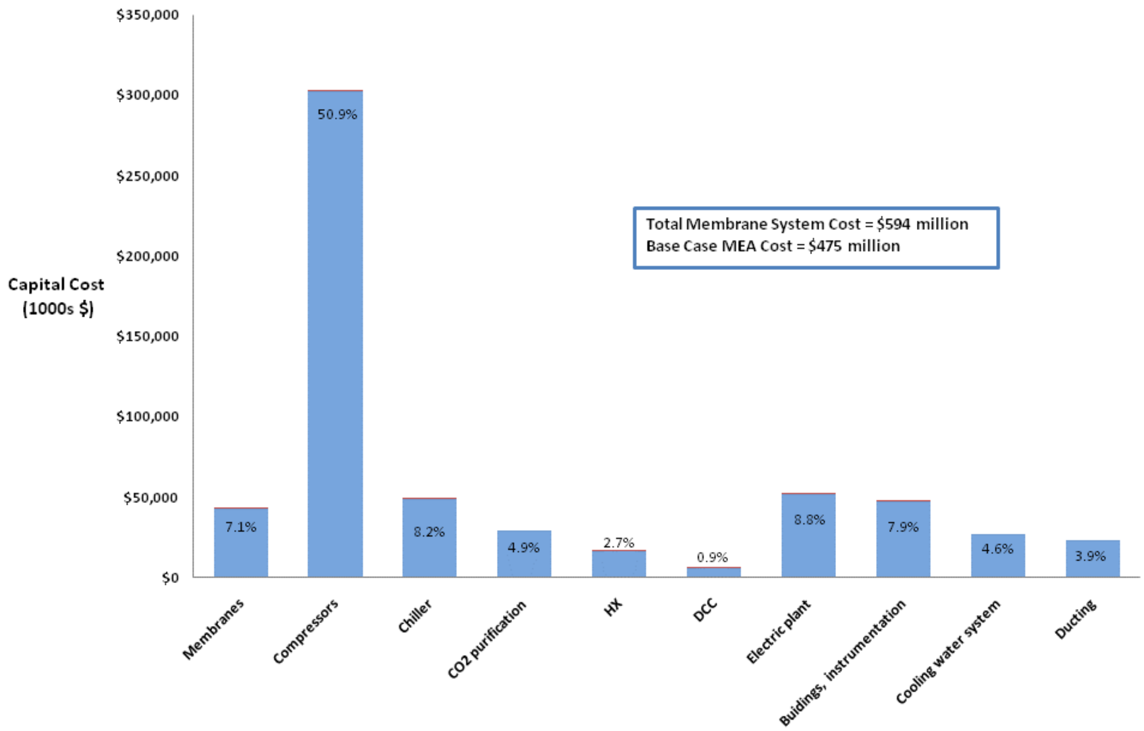


Figure 36. Breakdown of the estimated capital cost of the base case (2 bar feed) membrane system designed for 90% CO_2 capture from Conesville Unit 5. This estimate uses an average compression equipment installed cost of $1,640/kW and an installed membrane skid cost of $50/m^2. When 10% engineering fee and 20% project contingency are included, the total capital cost of the compression equipment is $2,150/kW and the membrane skids are $66/m^2.

One way to compare compression costs of various systems is to examine the cost of the rotating equipment normalized by the motor size (with units of $/kW). This is a rough measure of comparison because, in addition to motor size, there are many other factors that impact the cost of compression equipment. Other factors include the pressure ratio, the suction pressure and volumetric flow, the density of the gas being compressed, the materials of construction, and so forth. Nevertheless, for simplicity, compression costs in $/kW are used in the following sensitivity studies to examine the effect of these costs on membrane system competitiveness.

Table 17 shows that the average installed compression cost used in the EPRI/WP study for the base case membrane system is $1,640/kW. When a 10% engineering fee and 20% project contingency are included, the total capital cost of the compression equipment used in the study is $2,150/kW. This value is relatively high compared to compression costs obtained from other sources, as listed in Table 17. For example, in the 2010 revision of the DOE Bituminous Baseline report, CO_2 compression and drying costs are reported as $780/kW installed and $1,030/kW total capital cost (with fees and contingencies). Other cost estimates obtained by MTR and shown in Table 1 are also considerably lower than those used in the EPRI/WP study. The reason for these differences is not entirely clear. The compression equipment used in the base case MTR membrane design is mostly low-pressure-ratio compression or vacuum equipment, where the costs on a $/kW basis might be expected to be relatively high. Moreover, as described in the EPRI/WP Final Report, the EPRI/WP cost numbers are based on actual vendor quotes for today's equipment, so they cannot be dismissed. Nevertheless, based on the Table 1 data and cost improvements that might be expected to occur over the 10-20 year time horizon in which this technology would be fully commercialized, it seems reasonable to assume lower compression costs are possible for the membrane system.

Table 17. Compressor Equipment Costs for Various Large-Volume Gas Applications.

Application	Source	Compressor Type	Motor Power (kW)	Pressure Ratio	Inlet Pressure (bar)	Bare Erected Cost ($/kW)	Total Installed Cost[a] ($/kW)
Flue gas compression	EPRI/WP Final report (Section 5)	Centrifugal	--	2	1.1	1,650	2,150
Vacuum pumps and CO_2 compression	EPRI/WP Final report (Section 5)	Axial/ centrifugal	--	50	0.2	1,640	2,150
CO_2 compression	DOE Bituminous Coal Report	Multi-stage centrifugal	24,400	150	1	780	1,030
Flue gas compression	Vendor quotes to MTR	Axial	25,000	2	1.0	800	-
Syngas compression	Vendor quotes to MTR	5-Stage centrifugal	10,300	27	1.0	1,010	-

a. Assumes a 10% engineering fee, no process contingency, and 20% project contingency.

Figure 37 shows a breakdown of the total capital cost for the base case membrane system designed for 90% CO_2 capture from Conesville Unit 5, with compression costs equal to half the value of the EPRI/WP numbers. This amounts to a total installed compression equipment cost of $1,075/kW – a value similar to that used in the DOE Bituminous Baseline report for CO_2 compression and drying costs. The effect of this change on the total membrane system cost is dramatic. The membrane system swings from being about 25% more costly than the SOA amine system to being 10% less expensive. With these estimated costs, the membrane system would now have lower capital cost, lower operating cost and lower energy use compared to the amine process.

Examining the distribution of costs shown in Figure 37, compression equipment is still the most costly item for the membrane system, although it has been reduced from >50% of the total cost to about one-third of the plant cost. Compression optimization is a focus of CCS research, and future improvements in compressor efficiencies and cost reductions will clearly help the membrane system competitiveness. Other cost categories, besides the membranes, where cost reductions may be possible are in the electric plant and building and instrumentation. These two categories amount to about $100 million (23% of the total plant cost) in the EPRI/WP analysis, and with better plant integration of low-maintenance, easily-controlled membrane skids, can almost certainly be reduced.

The fact that compression is still the dominant cost for the base case membrane system, even with a lower-cost equipment assumption, suggests that other process designs that use no feed compression, or hybrid designs that use no feed compression or vacuum equipment, deserve more detailed analysis. Examination of a "no feed compression case" using advanced membranes is given later in this section. A preliminary discussion of hybrid designs with no vacuum equipment was given in Section 4. A more detailed study of these process configurations is recommended in future work.

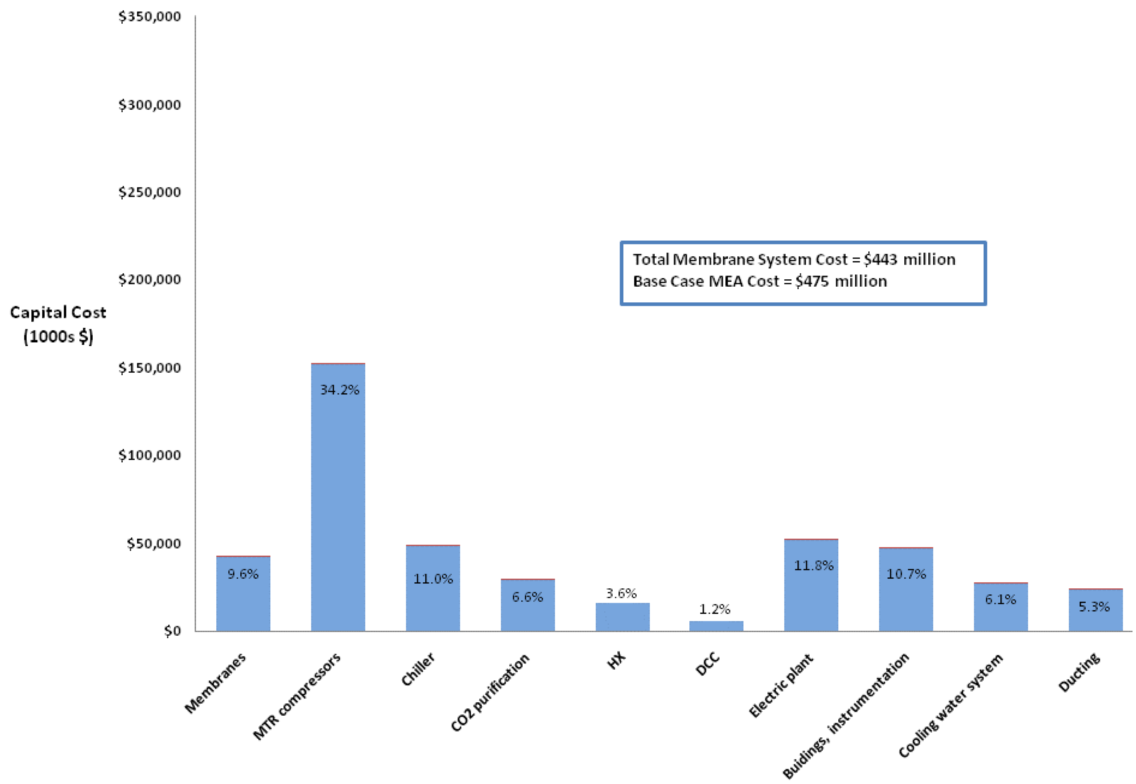


Figure 37. Breakdown of the estimated capital cost of the base case (2 bar feed) membrane system designed for 90% CO_2 capture from Conesville Unit 5. This estimate uses an average compression equipment installed cost of $820/kW and an installed membrane skid cost of $50/m^2. When 10% engineering fee and 20% project contingency are included, the total capital cost of the compression equipment is $1,075/kW and the membrane skids are $66/m^2.

Base Case Membrane Process Sensitivity Analysis

The following sensitivity study was conducted to more fully explore membrane system cost over a range of feed pressures, compression costs, and membrane performance. This analysis was limited to the base-case process shown in Figure 6 (Section 1) where a membrane CO_2 capture step operates in series with a sweep membrane step. Studies of other membrane process designs may yield different results, although we believe the central issues of how to generate affordable pressure ratio will apply generally to membrane-based flue gas CO_2 capture.

For the base-case Figure 6 design, a pressure ratio of 10 was chosen for the membrane CO_2 capture step, corresponding to a feed pressure of 2 bar and a permeate vacuum of 0.2 bar.

The vacuum pressure was selected based on a survey of equipment suppliers as to what would be the minimum pressure technically and economically feasible for an application of this scale.

The feed pressure was chosen based on prior MTR design studies that showed a pressure ratio of 10 was a good balance between compression costs (which increase with increasing pressure ratio) and membrane costs (which decrease with increasing pressure ratio).

In the following analysis, the feed pressure (and therefore, the pressure ratio) is changed to determine how this variable affects the overall system cost and incremental cost of electricity. Other variables examined include the compression equipment cost, membrane permeance and cost, and CO_2 capture percentage. Unless otherwise indicated, all other assumptions and calculation methods were identical to those applied in the Section 5 base case analysis.

Effect of Feed Pressure on Energy and Membrane Area Requirements

Partial pressure driving force is the means by which dense membranes separate gas species. The great challenge of using membranes for flue gas CO_2 capture is generating driving force for CO_2 separation in an affordable manner for a large flue gas stream that contains dilute CO_2 near atmospheric pressure. As described in Section 1, the base-case membrane design uses a combination of permeate vacuum and countercurrent air sweep to separate CO_2 while minimizing rotating equipment costs. If no feed compression is used, the pressure ratio on the CO_2 capture step is about 5 (1 bar feed / 0.2 bar vacuum permeate). This relatively low pressure ratio increases membrane area requirements and limits the purity of the captured CO_2 (which increases downstream CO_2 purification costs). For this reason, it is worthwhile to examine how feed pressure impacts the overall cost and performance of the base case design.

Figure 38 shows (a) the total energy used and (b) the membrane area required to capture 90% of the CO_2 from Conesville Unit 5 as a function of the pressure of the flue gas feed to the membrane process. The total energy used initially decreases with increasing feed pressure, shows a minimum around 2 bar, and then increases steadily at higher feed pressures. The reason for this somewhat complicated behavior can be explained as follows.

- At ambient pressure (~1.1 bar), while no energy is used for feed compression, the separation achieved by the CO_2 capture membrane step is poor because of the low pressure ratio. As a result, the CO_2-enriched permeate stream still contains significant amounts of nitrogen, and the energy required in the CO_2 compression and purification section of the process is high.

- As the feed pressure is increased to 2 bar, the purity of the captured CO_2 improves and the energy required for CO_2 compression/purification decreases more than the increase in energy required to compress the flue gas feed. Consequently, the total net energy used decreases.

- Above 2 bar, the improved CO_2 purity resulting from the increasing pressure ratio is no longer enough to compensate for the increased feed compression energy. The result is that the total CO_2 capture energy increases over the pressure range from 2 bar to the maximum pressure examined (15 bar).

Also shown in Figure 38(a) is the energy used by the base case MEA process to capture 90% of the CO_2 from Conesville Unit 5. Over a range of low feed pressures (1.5 to 3.5 bar), the membrane process uses less energy than the MEA process. Above 3.5 bar, the membrane process always uses more energy than the MEA process.

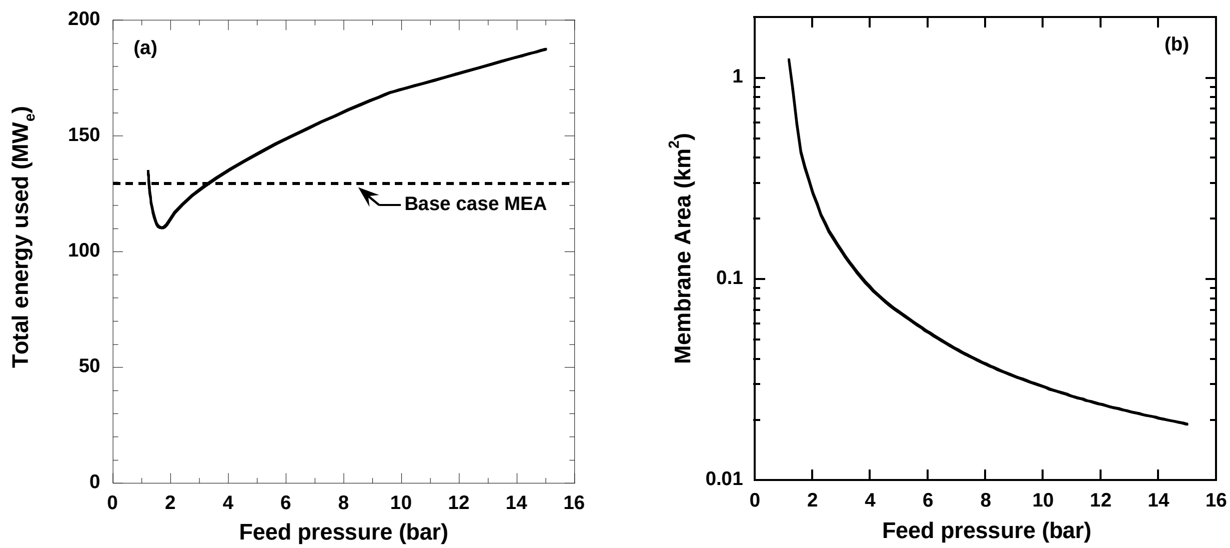


Figure 38. The (a) total energy used and (b) membrane area required for 90% CO_2 capture from Conesville Unit 5 as a function of the pressure of the flue gas feed to the membrane process. The total energy used includes CO_2 compression and purification, as well as feed compression to the value indicated on the x-axis. The base case membrane process (Figure 6, Section 1) with current membrane performance values was used in these calculations. The calculations shown in Figure 38 demonstrate why 2 bar feed pressure was chosen for the base case in the EPRI/WP analysis (see text for further discussion).

Figure 38(b) illustrates the effect of feed pressure on required membrane area. As expected, increasing the feed pressure dramatically reduces the required membrane area due to the increased driving force for CO_2 separation. This is especially true over the pressure range of 1 to 4 bar where the required membrane area drops by an order of magnitude. Clearly, higher feed pressure will reduce the membrane capital cost.

The calculations shown in Figure 38 demonstrate why 2 bar feed pressure was chosen for the base case in the EPRI/WP analysis. At 2 bar feed, the membrane area required is significantly lower than in the no feed compression case (1.1 bar). This means the membrane capital cost and footprint are lower at 2 bar feed compared to no feed compression. Moreover, with current membranes, 2 bar feed pressure appears to require the minimum total energy for CO_2 capture. Higher feed pressure will reduce the membrane area further, but the higher energy use will result in higher capital (compression equipment) and operating (electricity) costs.

Effect of Feed Pressure and Compression Cost on Capital Costs and LCOE

Figure 39 shows the effect of feed pressure and compression costs on the total capital cost of a membrane CO_2 capture process retrofitted to Conesville Unit 5. Three curves are shown in this figure, corresponding to compression equipment costs of \$500/kW, \$1,000/kW, and \$2,000/kW. Each of these curves has a shape similar to that described above in Figure 38(a) for energy use. As the feed pressure is increased, the total capital cost initially decreases to a minimum (between 1.5 and 3.5 bar), and then increases continuously at higher pressures. This cost increase with increasing feed pressure is most pronounced when the compression equipment costs are high (\$2,000/kW). As compression equipment becomes cheaper, the effect of increasing feed pressure on capital cost becomes less pronounced. In fact, at \$500/kW, the capital cost of a membrane CO_2 capture system is lower than that of the base case MEA system over the full range of feed pressures examined. In contrast, at \$2,000/kW – close to the value assumed for the membrane base case examined at 2 bar feed in Section 5 – the membrane system is always more expensive than the MEA system. *The breakeven point for the capital cost of membrane and MEA systems appears to be around \$1,000/kW.* In this case, the membrane process is cheaper at low feed pressure (between 1.5 and 3.5 bar), but becomes more expensive than the MEA system when higher feed pressure is used.

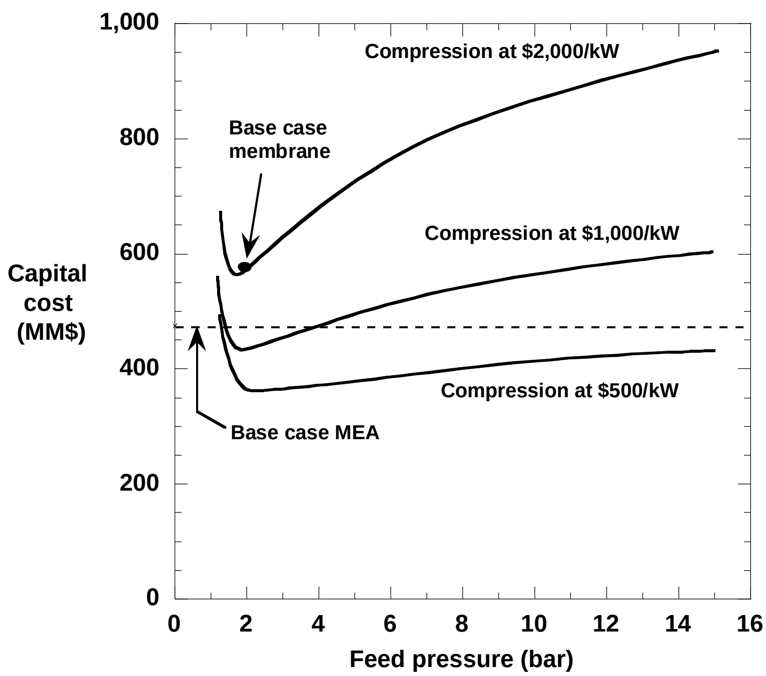


Figure 39. Effect of flue gas feed pressure and compression costs on the capital cost of a membrane system installed at Conesville to capture 90% of flue gas CO_2. These calculations assume membranes with a CO_2 permeance of 2,500 gpu at \$50/m^2 installed cost. The Section 5 study performed by EPRI/WP estimated the membrane compression costs would be in the range of \$2,000/kW; other external sources suggest compression costs closer to \$1,000/kW are more likely future values.

In addition to capital cost, another important metric when evaluating a retrofit CO_2 capture technology is the incremental levelized cost of electricity (LCOE). Figure 40 shows the incremental LCOE as a function of feed pressure and compression costs for a membrane CO_2 capture system. The shape of the curves is similar to that shown in Figure 40 for total capital cost. The incremental LCOE initially decreases to a minimum between 1.5 and 2.5 bar feed pressure and then increases continuously at higher feed pressures. Lower compression costs yield lower incremental LCOE values and reduce the dependence of LCOE on feed pressure. Also shown in Figure 40 is the incremental LCOE value for the base line MEA process.

At low pressure, all three of the membrane compression cost cases show potential to have lower LCOE than the MEA process. This reflects the fact that at low feed pressure, the membrane processes use less energy, have lower O&M costs, and for two cases ($500/kW and $1,000/kW) have lower capital costs than the MEA process. In contrast, at high feed pressures (>10 bar), all of the membrane compression cost cases have higher LCOE than the MEA process. At these high pressures, the higher energy use of the membrane processes overwhelms the lower O&M costs and lower capital costs (for $500/kW) of the membrane systems.

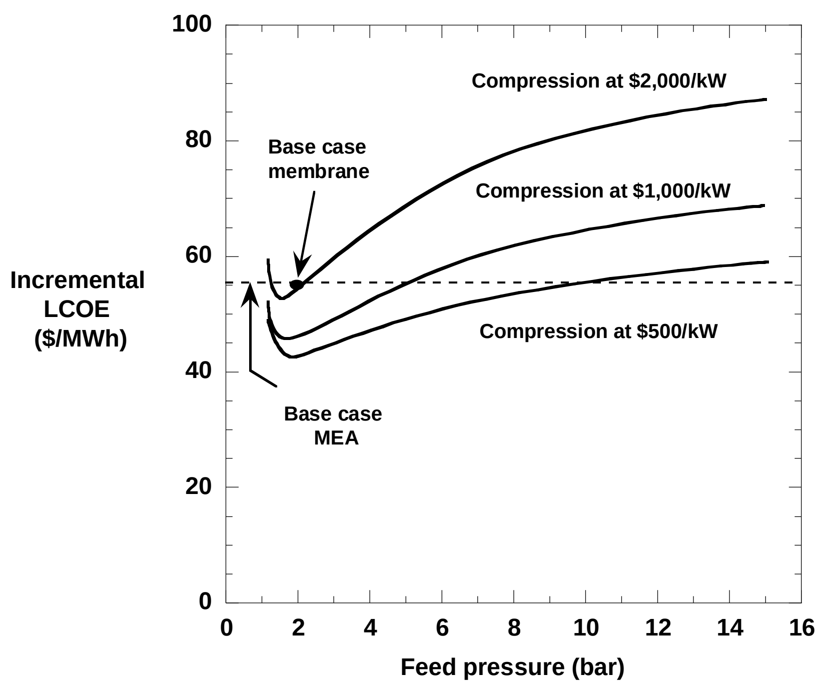


Figure 40. Effect of flue gas feed pressure and compression costs on the incremental LCOE for a membrane system installed at Conesville to capture 90% of flue gas CO_2. The shape of the curves is similar to that shown in Figure 39 for total capital cost. At low pressure, all three of the membrane compression cost cases show potential to have lower LCOE than the MEA process.

In summary, Figures 38-40 show that for the base case membrane process design:

- Low feed pressures are preferred to minimize energy use and compression costs.

- Both the total capital cost and the incremental LCOE for a membrane CO_2 capture system show similar behaviors, with minimums occurring between 1.5 and 3 bar feed pressure.
- At the present time, the cost of compression equipment is the single most influential parameter for determining costs of a membrane CO_2 capture system. This is in large part due to the broad range in assumptions (Δ of as much as 300%) than can be made for costs of compression equipment.
  - At high estimated compression equipment costs of $2,000/kW, the base case membrane system will likely have higher capital cost than the base case MEA system and approximately equivalent incremental LCOE values (because the membrane process uses less energy and has lower O&M costs). If lower compression costs are not possible, different membrane process designs that minimize required compression equipment will be favored.
  - At lower compression equipment costs, the membrane system can have lower capital costs as well as lower incremental LCOE compared to the base case MEA system.

Effect of Membrane Performance

We also conducted a sensitivity analysis to examine the impact of membrane permeance-normalized cost on system performance. Permeance-normalized cost is a parameter – here expressed in \$/(m^{2}gpu) – that combines the effects of membrane permeance with the cost required to produce that membrane. The permeance-normalized cost can be minimized by either increasing the permeance of the membrane or reducing the production cost of the membrane. Based on the performance of today's best membranes (2,500 gpu) and the expected cost of the membrane module skid for a large plant at US\$50/m^2, our target permeance-normalized cost is about \$0.02-0.03/(m$^{2\cdot}$gpu). In the EPRI/WP study, a cost number of \$0.1/(m^{2}gpu) was used.

Figure 41 shows the optimum feed pressure (identified as the pressure providing the lowest incremental LCOE in the Figure 40 plots) for different membrane permeance-normalized costs and compression equipment costs. This figure illustrates a clear trend that can be summarized as follows:

- As the permeance-normalized membrane cost is reduced, the feed pressure at the minimum incremental LCOE decreases. Said another way, as the membrane becomes very permeable and/or low cost, less feed compression is preferred. In fact, *when the permeance-normalized cost reaches \$0.01/(m^2 gpu) or less, almost no feed compression (<2 bar) should be used, regardless of how cheap or expensive compression equipment is*.
- When membranes have lower permeance or are more costly [> \$0.1/(m^2 gpu)], and compression equipment is relatively inexpensive, there is a better argument for using more feed compression. For example, if compression equipment costs \$500/kW and membrane permeance-normalized cost is \$0.1/(m^2 gpu), the lowest incremental LCOE will be achieved when the flue gas feed to the membrane system is compressed to ~5 bar.

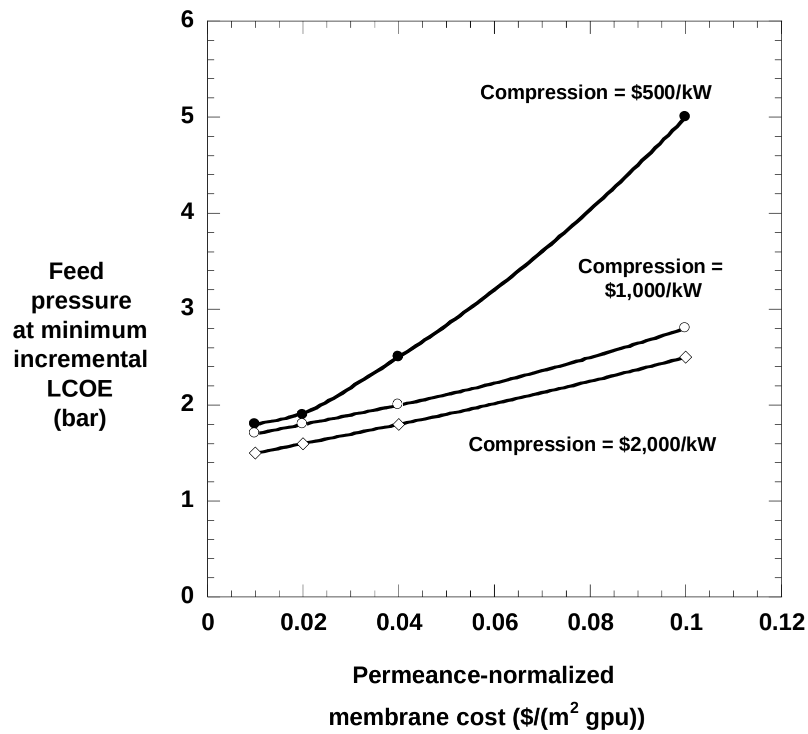


Figure 41. Optimum feed pressure at the minimum LCOE as a function of the permeance-normalized membrane cost and compression equipment costs for a membrane system installed at Conesville to capture 90% of flue gas CO_2. The current MTR target for permeance-normalized membrane cost is about $0.0.02-0.03/($m^2$ gpu).

Effect of CO_2 Capture Rate

Another variable of interest for a capture technology is the cost of CO_2 capture as a function of the fraction of CO_2 generated by a power plant that is captured. Currently, DOE and the CCS community are focused on 90% CO_2 capture from power plants as a baseline emissions mitigation target. However, initial CO_2 emission regulations, such as AB32 passed in California, are likely to impose penalties on emissions above that of a natural gas-fired power plant. Because natural gas (mostly methane) has a low carbon/hydrogen ratio compared to coal or other fossil fuels, combustion of this gas produces less CO_2 per unit of power generated. For example, a natural gas-fired power plant produces about half the CO_2 emissions per MW of electricity generated compared to an average coal-fired plant. For a coal plant to match the emissions intensity (in mass CO_2 emitted/MW produced) of a gas plant, the coal plant will need to capture about 50% of its CO_2 emissions. Consequently, it is of interest to examine the impact of the CO_2 capture rate on the cost of CO_2 capture.

Previous studies have shown that the total cost of an amine capture system will decrease as the fraction of CO_2 captured is reduced. This will be the case for a membrane system as well. For example, if 50% capture is required rather than 90%, a smaller membrane system can be used to treat a portion of the flue gas. Another portion of the flue gas stream can bypass the membrane system and go directly to the stack. One potential advantage of a membrane process in such a regulatory environment is the flexible, modular nature of the technology. If different CO_2

capture rates are required over the life of the plant (for example, a 50% capture rate that is later increased to 90% to adjust to progressive regulations), additional membrane modules can simply be added when needed.

Figure 42 shows the CO_2 capture (or mitigation) cost per ton of CO_2 as a function of the CO_2 capture rate. This normalized capture cost (in $/ton CO_2 captured) demonstrates the relative efficiency of a capture process at different capture rates. For a 2 bar feed pressure membrane case, the cost of capture per ton of CO_2 decreases with increasing capture rates from 50-95%. This result indicates that the membrane system is operating most cost-effectively at high CO_2 capture rates (90-95%). At lower capture rates, the system does not capture enough CO_2 to compensate for the capital cost of the system, and therefore, the capture cost per ton of CO_2 is high. Although not shown, at very high capture rates (>95%), the capture cost per ton CO_2 increases rapidly because of large energy and membrane area increases associated with removing small incremental amounts of CO_2 from the flue gas. These trends are similar for the different compression equipment cost curves illustrated in Figure 42, with the logical result that lower compression costs yield lower CO_2 capture costs. For comparison, the CO_2 capture cost of the base case MEA system is also shown in Figure 42.

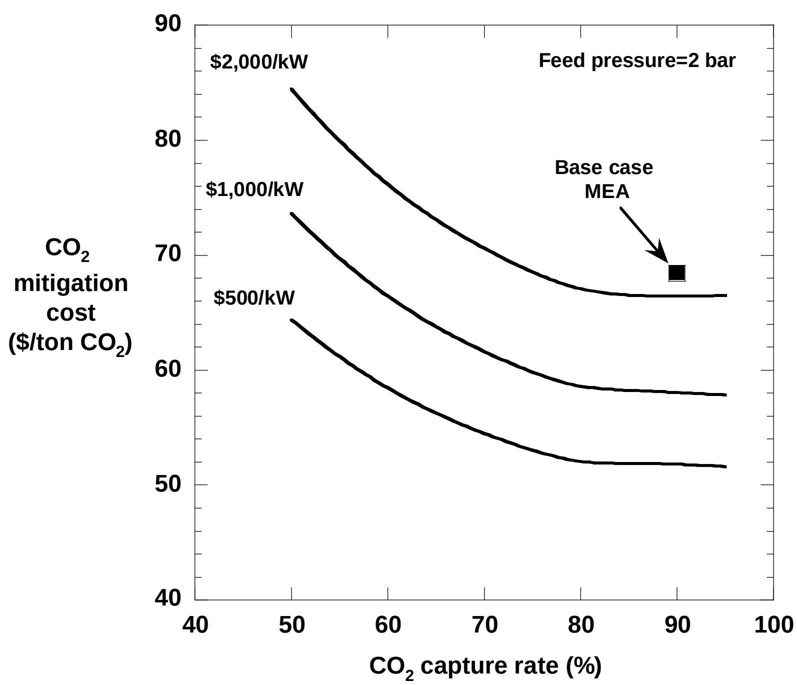


Figure 42. Normalized CO_2 capture (mitigation) cost as a function of CO_2 capture rate for the base case membrane system with 2 bar feed pressure.

Figure 43 shows the normalized cost of CO_2 capture as a function of the CO_2 capture rate for a membrane system with minimal feed compression (1.2 bar feed pressure). The shape of the different compression equipment cost curves for this 1.2 bar system is different from those shown in Figure 42 for the 2 bar feed case. With 1.2 bar feed pressure, the capture cost per ton of CO_2 shows a minimum around 70% CO_2 capture. This result can be interpreted as follows:

- At low capture rates (<70%), the capture cost per ton of CO_2 falls with increasing capture rate for the same reason as that mentioned above for the 2 bar case – not enough CO_2 is captured to compensate for the capital cost of the capture equipment.
- Above 70% capture, the capture cost per ton of CO_2 increases rapidly because the low feed pressure (and small driving force) means that the membrane area must be increased dramatically to make incremental gains in CO_2 capture rate.

The consequence of these trends is that different capture rates may call for different membrane system designs. Lower capture rates of 50-70% are better suited for minimum feed compression, while a 90% capture system operates most cost-effectively with higher feed pressure (2 bar).

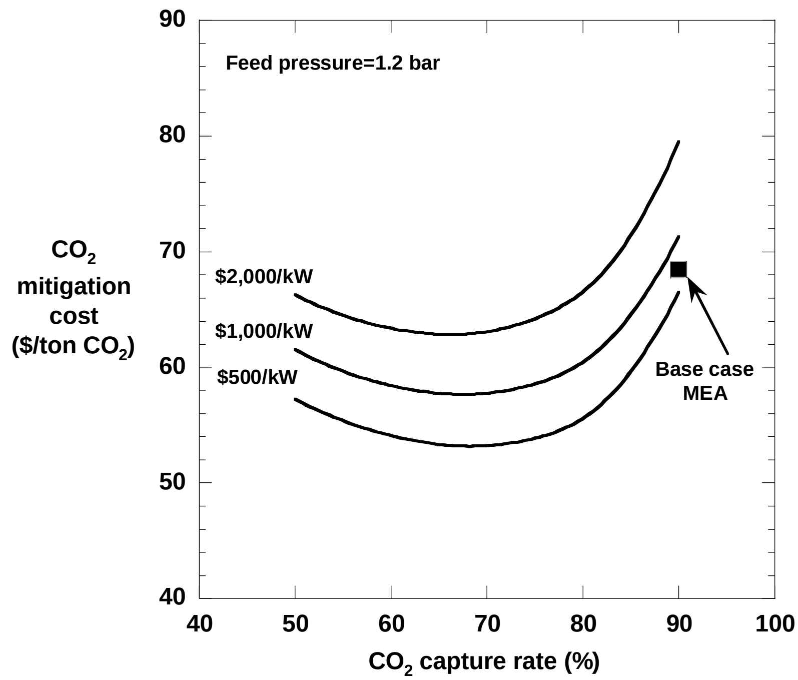


Figure 43. Normalized CO_2 capture (mitigation) cost as a function of CO_2 capture rate for the base case membrane system with 1.2 bar feed pressure.

The calculations shown in Figures 42 and 43 demonstrate that a membrane CO_2 capture system may offer advantages not highlighted in the EPRI/WP base case analysis (in this case, lower CO_2 capture costs at lower CO_2 capture rates). Other factors that were not quantified in the base case analysis, and that may offer membranes a competitive advantage, are summarized at the end of this section.

Effect of Using Advanced Membranes and No Feed Compression

The Section 5 cost analysis is based on current membrane and rotating equipment performance and costs. The sensitivity analysis described above indicates that compression costs are the most important cost item for a membrane CO_2 capture system. This result suggests that using no feed compression should be a way to minimize membrane CO_2 capture systems costs. However, Figures 38-40 show that with current membranes the increase in membrane area and reduced permeate CO_2 purity at no feed compression make this a more expensive option. As a final consideration in this sensitivity study, we examine the case where no feed compression is combined with advanced membranes to produce an optimized capture system. The following assumptions are made for this analysis:

1. Membranes with CO_2 permeances of 5,000 gpu and CO_2/N_2 selectivities of 60 at 50°C can be developed. Although not available today, the likelihood of membranes with these properties being developed in a 5- to 10-year time frame is high.
2. Low-cost rotating equipment ($500/kW) with high efficiency (93%) is available. This appears likely for the CO_2 compression equipment because optimization of this equipment is a CCS research priority. The availability of high-performance, low-cost vacuum equipment is more uncertain.
3. Boiler efficiency is not affected by CO_2 recycle and make-up air to replace O_2 lost in the sweep module is not required as long as there is stoichiometric air in the boiler. Initial feedback on this assumption looks promising, but more study is needed.

Table 18 summarizes the itemized capital cost of the advanced membrane system retrofitted to Conesville Unit 5. Table 19 shows the breakdown of the parasitic energy load for 90% CO_2 capture. With advanced technologies, the overall capital cost of the membrane system is $252 million, significantly lower than the base case MEA system ($475 million). The parasitic energy load for 90% CO_2 capture with the advanced membrane system is approximately 96 MW_e. This is approximately 25% lower than the base case MEA process (130 MW_e).

Table 18. Itemized System Costs Using Next Generation Membrane and Equipment Assumptions.

Equipment	US$ Thousands
Membrane modules	65,040
Booster air fan	1,799
Flue gas blower	5,939
Vacuum pumps and CO_2 compressors	33,099
CO_2 chilling system	49,476
CO2 purification system	29,268
Cooling water system	12,287
Recycle gas turbo expander	1,699
Duct and construction	53,561
Total plant cost	252,168

Table 19. Breakdown of Parasitic Energy Load for 90% CO_2 Capture by the Membrane System, Using Next Generation Membrane and Equipment Assumptions.

Equipment	Parasitic Load
Booster air fan	3.53
Flue gas blower	11.64
Vacuum pumps	29.48
CO_2 compressors	33.76
CO_2 dryer	0.13
Chiller compressors	13.25
CO_2 pump	2.28
Auxiliary cooling service	1.52
Additional plant auxiliary loads	3.73
Recycle gas turbo expander	-3.37
Total auxiliary load	95.95

Figure 44 provides a comparison of the incremental LCOE costs for several MEA and membrane technology options. The incremental LCOE for CO_2 capture using the advanced membrane system with low-cost compression technology is \$33.3/MWh, 40% lower than the advanced MEA-1A process (\$53.0/MWh).

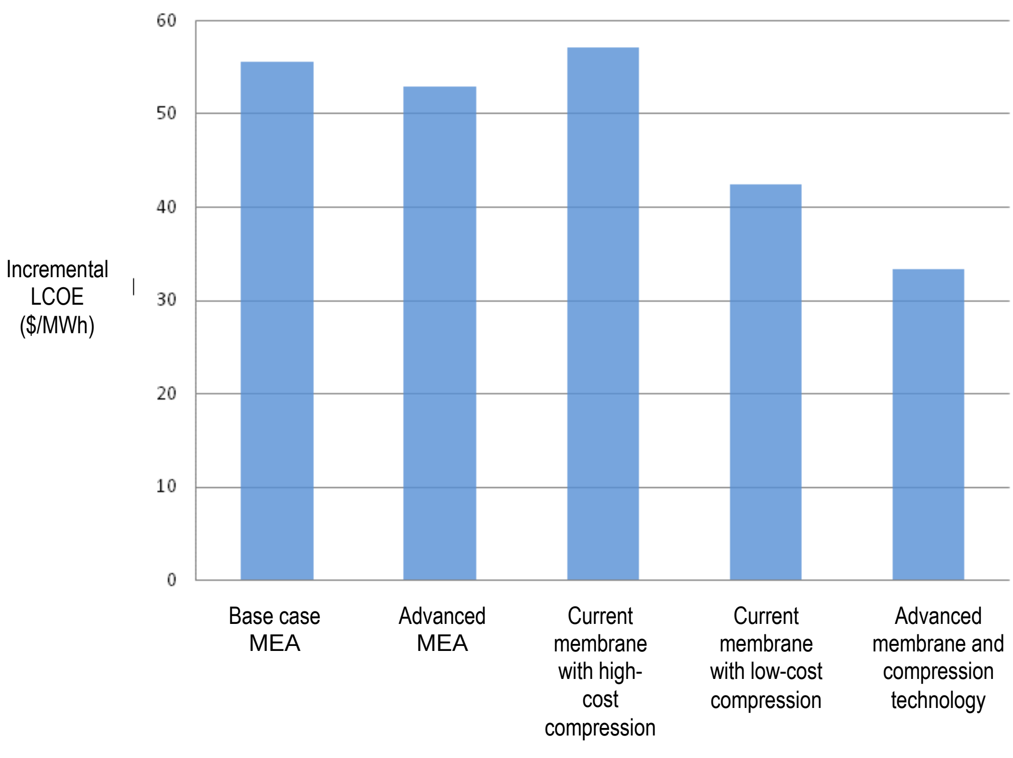


Figure 44. Incremental LCOE costs comparisons for base case and advanced amine and membrane technology options. With advanced technologies, the incremental LCOE for CO_2 capture by the membrane system is 40% lower than the state-of-art MEA process.

Figure 45 is an updated version of Figure 40 showing the effect of feed pressure on the incremental LCOE for the advanced membrane technology case. For comparison, the base case design with current membranes (studied in Section 5) with compression at $2,000/kW (current) and $500/kW (future) are also shown in this figure. As discussed earlier, reducing the cost of compression equipment alone dramatically reduces the incremental LCOE for the membrane capture process. For example, the incremental LCOE at 2 bar feed pressure can be reduced from $57.1/MWh with $2,000/kW compression equipment to $43/MWh with $500/kW compressors. Improving the membrane properties further reduces the incremental LCOE, particularly at low feed pressure. In fact, with advanced membranes, feed compression equipment can be omitted completely and the incremental LCOE can be reduced to as low as $33.3/MWh.

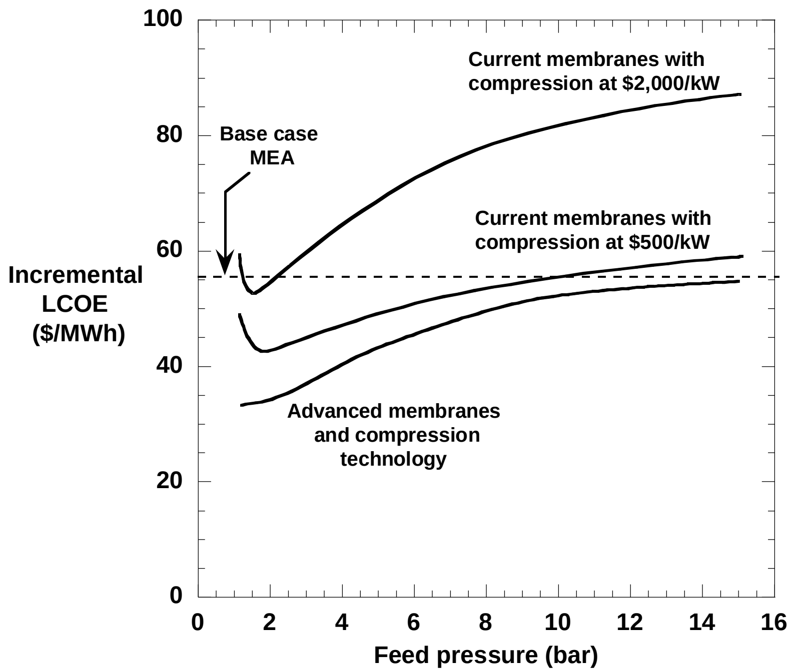


Figure 45. Incremental LCOE costs as a function of feed pressure and compression costs using the advanced membrane technology assumptions. With advanced technologies, the incremental LCOE for CO_2 capture by the membrane system is as much as 40% lower than for the MEA process.

Other Competitive Issues

As discussed earlier, there are other competitive issues that were not quantified in this report, but that may offer advantages for membrane systems applied to post-combustion CO_2 capture. Table 20 summarizes a number of these factors; it is recommended that future studies examine these issues in more detail.

Table 20. Summary of Other Factors that Impact Capture Technology Competitiveness

Issue	Potential Additional Advantages Offered by Membranes
Lower CO_2 capture levels, different product specifications	Presently, there is considerable debate over the real requirements for the CO_2 product specification (WP, p.15 and 73 of Report) and the CO_2 capture rate. The current focus on 90% capture at near 100% purity generally favors capture by absorption processes. However, these targets may exceed sequestration, market, and initial regulatory requirements. Membranes gain significantly compared to amines at capture levels of 70% or lower (See Figure 41, Section 6). Similarly, lower CO_2 purity requirements would favor a membrane process, but not an amine absorption unit. For example, in EOR applications where N_2 is often injected with CO_2, the cryogenic distillation column used in the EPRI/WP base case membrane design to purify CO_2 to >98% would not be required.
High-sulfur feeds	Membranes are stable in high-sulfur environments and will co-capture SO_2 with CO_2. In contrast, amines require deep FGD to prevent heat stable salt formation. The analysis conducted in Section 5 gives no credit to membranes for savings in FGD limestone that would be possible if the membrane system is allowed to co-capture SO_2. Particularly for power plants where high-sulfur coals are used and/or an FGD system is not currently in place, a membrane system could be a low cost means to meet SO_2 and CO_2 emissions limits. The best result would be to sequester the co-captured SO_2 with the CO_2. If this is not possible, SO_2 can be removed at modest cost from the small high-pressure CO_2 stream.
Process emissions and EPA/local permitting	As a passive separation process, membranes do not emit VOCs or require hazardous chemicals handling or disposal. This may make permitting issues related to installation of the membrane capture system much easier compared to amine absorption. Amine systems require daily handling and disposal of large amounts of make-up amine solution, and add concerns about emissions of nitrosamine and nitramide by-products in the flue gas.
Natural gas-fired power generation	The low capital cost of natural gas power plants combined with an abundance of low-cost domestic natural gas favor expanded use of NGCC power plants in the coming decades. Eventually, CO_2 capture from these plants will become an issue as well, and current amine systems are ill-suited for this application. Current estimates show that the cost of CO_2 capture from dilute NGCC flue gas by amine processes will be as much as 40% more than from a coal plant on a per-ton-of-CO_2 basis. In contrast, as described in Section 4, membranes may be able to capture CO_2 from NGCC flue gas at a lower cost than from coal plants because of the large amount of excess air used in a combustion turbine, which allows more CO_2 to be captured by energy-efficient sweep recycle.
Possible changing regulatory environment	The modular nature of membrane separation units are a potential advantage if different CO_2 capture rates are required over the life of the plant (for example, a 50% capture rate that is later increased to 90% to adjust to progressive regulations). In this scenario, additional membrane modules can simply be added when needed.
Grassroots vs. retrofit	As is probably true for most capture technologies, a membrane system could be designed to operate more efficiently at a new plant compared to a retrofit. In particular, lower capture costs could be realized by optimizing boiler design with CO_2 sweep recycle.

7. CONCLUSIONS AND FUTURE WORK

This DOE NETL-funded project examined the potential of an energy-efficient membrane process to capture CO_2 from power plant flue gas. The work was conducted by the project partners from October 1, 2008 through March 31, 2011.

The primary goal of this research program was to demonstrate, in a field test, the ability of a membrane process to capture up to 90% of CO_2 in coal-fired flue gas, and to evaluate the potential of a full-scale version of the process to perform this separation with less than a 35% increase in the levelized cost of electricity (LCOE). Membrane Technology and Research (MTR) conducted this project in collaboration with Arizona Public Services (APS), who hosted a membrane field test at their Cholla coal-fired power plant, and the Electric Power Research Institute (EPRI) and WorleyParsons (WP), who performed a comparative cost analysis of the proposed membrane CO_2 capture process.

In addition to a field demonstration and a systems cost analysis, other project activities included membrane and module development, and process design optimization. Key results from this program are summarized below.

- The Polaris CO_2 capture membrane was further optimized. During this project, the CO_2 permeance of Polaris membrane was doubled while maintaining the CO_2/N_2 selectivity. This is an important accomplishment because doubling of CO_2 permeance halves the skid cost and footprint.
- The stability of Polaris membranes in a flue gas environment containing SO_2 was confirmed. Laboratory tests showed no degradation in Polaris membrane performance during two months of continuous operation with simulated flue gas containing up to 1,000 ppm SO_2.
- A slipstream field test at the APS Cholla power plant was conducted with commercial-size Polaris modules. This system ran for three months during summer 2010 and demonstrated stable module performance, as well as successful countercurrent operation with air as a sweep stream. This field test is the first demonstration of commercial-sized membrane modules treating actual coal-fired power plant flue gas.
- Process design studies confirmed that selective recycle of CO_2 using a countercurrent membrane module with air as a sweep stream lowers the minimum energy required for CO_2 capture. The sweep membrane unit can double the concentration of CO_2 in coal flue gas with almost no energy input. This pre-concentration of CO_2 by the sweep membrane reduces the minimum energy of CO_2 separation by up to 40% for coal flue gas.
- A variation of the selective recycle membrane design may be even more promising for CO_2 capture from NGCC flue gas where the CO_2 concentration can be increased from 4% to 20% with little energy input.
- An analysis of the base case membrane CO_2 capture process by EPRI and WP suggests that CO_2 recycle to the coal boiler is feasible with minimal impact on boiler performance; however, further study by a boiler OEM is recommended.
- For a base case membrane process using a combination of slight feed compression (2.0 bar), permeate vacuum (0.2 bar) and current compression equipment costs ($2,150/kW), the membrane capture process can be competitive with the base case MEA process at 90% CO_2 capture from a coal-fired power plant. The EPRI/WP analysis indicates the membrane process will cost about 25% more than the MEA process (with 30% uncertainty), but will use less energy and have lower O&M costs. The overall effect is that the incremental LCOE for the base case membrane process is equal to that of a base case MEA process ($56/MWh), within the uncertainty in the analysis.

- For the base case membrane system with compression equipment priced at \$2,150/kW, compression costs make up more than 50% of the membrane CO_2 capture system costs. Lower compression costs, such as those used in the recent DOE Bituminous Baseline report (\$1,030/kW), will dramatically improve the competitiveness of a membrane process. For example, low-cost compression equipment will lower the membrane incremental LCOE to \$43/MWh. It seems likely that in the time horizon predicted for commercialization of CO_2 capture technology (ready by 2020), lower compression costs can be expected.
- Even with low-cost compression, it appears unlikely that high feed compression (>5 bar) membrane systems can be competitive with amine absorption due to the capital cost and energy consumption of this equipment. Similarly, low vacuum pressure (<0.2 bar) cannot be used due to poor efficiency and high cost of this equipment.
- High membrane permeance is important to reduce the capital cost and footprint of the membrane unit. CO_2/N_2 selectivity is less important because it is too costly to generate a pressure ratio where high selectivity can be useful. We recommend that membrane targets be set at 5,000 gpu for CO_2 permeance and 50 for CO_2/N_2 selectivity. Advanced membranes with these properties, operating with no feed compression and low-cost CO_2 compression equipment, would produce an incremental LCOE of \$33/MWh at 90% capture (40% lower than the advanced MEA case).
- A potential sweet spot exists for membranes if 50-70% CO_2 capture is acceptable. There is a minimum for membranes in the cost of CO_2 avoided/ton at 60% CO_2 capture that is 20% lower than at 90% capture.

Based on the findings in this project, it appears that the biggest hurdle to use of membranes for post-combustion CO_2 capture is the current cost of auxiliary compression equipment. An all-membrane capture process must use some flue gas feed compression and/or permeate vacuum to generate a driving force for CO_2 separation. There are cost and availability uncertainties associated with the large feed compressors and permeate vacuum pumps that would be required for an all-membrane capture process. Because compression is a focus of CCS research, it seems likely that more-efficient, more-affordable compression equipment will be available when CO_2 capture from power plants is eventually commercialized. However, this may not be the case for vacuum equipment because it is not required by other capture technologies (such as amines). For this reason, we recommend that future membrane project work focus on rotating equipment options and development needs.

An interesting approach to reduced membrane system reliance on compression equipment is to use sweep membranes in parallel with another CO_2 capture technology that does not require feed compression or vacuum equipment (such as amines). Hybrid designs that utilize sweep membranes for selective CO_2 recycle show potential to significantly reduce the minimum energy of CO_2 separation. This is particularly true for NGCC flue gas where sweep membranes can pre-concentrate CO_2 from 4% to 20% with little energy input. We believe such hybrid designs warrant further examination.

A key to the viability of the membrane selective recycle design (for coal or natural gas) is the assumption that CO_2 recycled to the combustion process does not adversely impact the combustion efficiency. A preliminary evaluation by EPRI/WP in this report suggests that it is possible to recycle CO_2 to a coal boiler without significantly affecting the efficiency of the boiler and steam cycle. Future work should seek to confirm this finding, preferably through testing on a small-scale boiler. This work should also explore the possibility of optimizing boiler design to benefit from the increased mass flow associated with CO_2 recycle.

Another area where future work can reduce the cost of CO_2 capture with membranes is improved module design. Low pressure drop through modules is crucial to reduce blower power requirements and high packing density is needed to reduce system footprint. The low-pressure, high-volume flow rate of flue gas is different from conventional gas streams treated by membranes and will require specialized modules. In addition to enhancing performance, quantifying module lifetime will be important to making realistic system cost estimates. This lifetime information can only be obtained from slipstream tests with real flue gas. For this reason, it is critical that slipstream tests, like those described in this report, continue to be used as a field laboratory to study membrane, module, and equipment reliability.

Finally, membrane systems have a number of unique attributes that may be beneficial for post-combustion CO_2 capture. Examples include:

- Membranes are not poisoned by SO_2, and in fact will remove SO_2 from flue gas even more efficiently than they will remove CO_2. This suggests the possibility of co-capture of SO_2 and CO_2 by membranes. If co-capture is viable, cost savings associated with replacing two unit operations (FGD for SO_2 removal and a CO_2 capture technology) with a single membrane system could be realized.
- As a passive separation process, membranes do not emit VOCs or require hazardous chemicals handling or disposal. This may make permitting issues related to installation of the membrane capture system much easier compared to amine absorption. Amine systems require daily handling and disposal of large amounts of make-up amine solution, and must address additional concerns about emissions of nitrosamines and nitramides as by-products in the flue gas.
- The modular nature of membrane separation units are a potential advantage if different CO_2 capture rates are required over the life of the plant (for example, a 50% capture rate that is later increased to 90% to adjust to progressive regulations). In this scenario, additional membrane modules can simply be added when needed.

These potential benefits and their impact on the competitiveness of a membrane CO_2 capture system have not been quantified in this report. We recommend that future work examine these items in more detail to clarify their importance in a comparative technology assessment.

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9. ACKNOWLEDGEMENTS

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10. EPRI/WP APPENDIX

Documents submitted to MTR by Electric Power Research Institute (EPRI) and WorleyParson Group Inc. (WP) for work conducted on subcontracts for DOE NETL contract DE-NT0005312 can be found in the appendix that follows.

EPRI/WP Appendix

Documents submitted to MTR by

Electric Power Research Institute (EPRI) and

WorleyParsons Group Inc. (WP)

for work conducted on subcontracts for

DOE NETL contract DE-NT0005312

Major contents include

- EPRI Cover Letter and LCOE Analysis
- WP Final Report for EPRI and MTR
  (called the EPRI/WP Final Report)

EPRI | ELECTRIC POWER RESEARCH INSTITUTE

March 22, 2011

Tim Merkel, Ph.D.
Director of Process Research and Development
Membrane Technology & Research
1360 Willow Road
Menlo Park, CA 94025

Subject: Final Report on DOE-NETL Project "Membrane Process for CO_2 From Flue Gas"

Dear Dr. Merkel,

This report comprises EPRI's final report for the subject contract. Two documents are attached:

1) *Cost Analysis*, including Appendix A. Shows EPRI's LCOE Analysis of the MTR process.

2) *Evaluation of MTR's CO_2 Membrane Process for Capturing CO_2 from Power Plant Flue Gas*. Shows WorleyParson's analysis of the MTR process.

Please let me know if you have any questions. I may be reached at 650-855-2383 or abhown@epri.com.

Regards,

Abhoyjit Bhown

Abhoyjit S. Bhown, Ph.D.
Sr. Project Manager

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COST ANALYSIS

LEVELIZED COST OF ELECTRICITY ANALYSIS AND RESULTS

An economic analysis was performed comparing the MTR technology with the MEA-1 process. The levelized cost of electricity (LCOE) and the CO_2 mitigation cost were evaluated to quantify the impact that the different CO_2 capture technologies has on the cost of electricity. Both are reported as the incremental cost of CO_2 capture compared to the base case Conesville 5 unit. The LCOE calculations followed the same methodology as was used in the 2007 DOE/NETL report "Carbon Dioxide Capture from Existing Coal-Fired Power Plants" [1], where LCOE is defined as the levelized annual capital charge plus the levelized annual operating costs, or:

$$LCOE_P = \frac{(CCF_P)(TIC) + [(LF_{F1})(OC_{F1}) + (LF_{F2})(OC_{F2}) + \cdots] + (CF)[(LF_{V1})(OC_{V1}) + (LF_{V2})(OC_{V2}) + \cdots]}{(CF)(MWh)}$$

where:

LCOE = levelized cost of electricity over P years
P = levelization period (e.g., 10, 20, or 30 years)
CCF = capital charge factor for a levelization period of P years
TIC = total investment cost [the sum of bare erected costs (includes costs of process equipment, supporting facilities, direct and indirect labor), detailed design costs, construction/project management costs, project contingency, process contingency, and technology fees]
LF_{Fn} = levelization factor for category n *fixed* operating cost
OC_{Fn} = category n *fixed* operating cost for the initial year of operation (but expressed in "first-year-of-construction" year dollars)
CF = capacity factor
LF_{Vn} = levelization factor for category n *variable* operating cost
OC_{Vn} = category n *variable* operating cost at 100% capacity factor for the initial year of operation (but expressed in "first-year-of-construction" year dollars)
MWh = annual net megawatt-hours of power generated at 100% capacity factor

Because the cost of make-up power to account for the reduction in net electricity production is included in the LCOE calculations, the annual net kilowatt-hours for the plant is assumed to be the same as the base case.
The CO_2 mitigation cost is defined in the 2007 DOE/NETL study as:

$$CO_2\ Mitigation\ Cost = \frac{(LCOE_{Cp} - LCOE_{Ref})}{(CO_{2\,Ref\ emitted} - CO_{2\,Cp\ emitted})}$$

where:

CO_2 Mitigation Cost = \$/tonne of CO_2 avoided
CO_2 = carbon dioxide (tonnes/MWh at plant capacity factor)
LCOE = levelized cost of electricity (\$/MWh)
$_{Cp}$ = capture plant
$_{Ref}$ = reference plant

Because the LCOE for this study is reported as the incremental cost of the retrofit, it is therefore already the equivalent of ($LCOE_{Cp}$ – $LCOE_{Ref}$).
The economic assumptions, as used in the 2007 DOE/NETL study, and well as the feedstock prices, are shown in Exhibit 1.

Exhibit 1
Economic assumptions

Capital Charge Factor	0.175
O&M Levelization Factor	1.1568
Feedstock Levelization Factor	1.1651
Coal Cost ($/MMBtu)	1.80
Natural Gas Cost ($/MMBtu)	5.95

Because the cost of electricity used to calculate the make-up power cost is already a levelized cost, a levelization factor is not applied to the make-up power.

Three different LCOE and CO_2 mitigation cost scenarios were analyzed for the MTR and MEA cases: using the base assumptions used in the 2007 DOE/NETL study, adding a nominal cost for CO_2 transportation and storage, and including the emissions associated with make-up power in the CO_2 mitigation cost analysis.

LCOE and CO_2 Mitigation Cost – Base Assumptions

Exhibit 2 compares the LCOE and CO_2 mitigation costs of the MTR case and the MEA cases. The LCOE is broken down into the capital cost for the retrofit, the incremental fixed and variable O&M costs, incremental coal and natural gas costs, and the cost of make-up power from the reduction in net electricity production.

Exhibit 2
LCOE and CO_2 Mitigation Cost – DOE/NETL Base Assumptions

	MTR	MEA-1	MEA-1a
Capital	32.2	25.7	25.7
Fixed O&M	1.7	0.9	0.9
Variable O&M	2.9	7.4	7.4

Coal	0.2	0.0	0.0
Natural Gas	0.0	0.2	0.2
Make-Up Power	20.2	21.4	18.7
Total LCOE ($/MWh)	**57.1**	**55.6**	**53.0**
CO_2 Mitigation Cost ($/tonne)	**70.4**	**68.4**	**65.2**

LCOE and CO_2 Mitigation Cost – Added CO_2 T&S Cost

When CO_2 is captured at a plant, it must be transported in a pipeline and injected in the ground for storage. While the plant cost boundary ends at the CO_2 compression stage and does not include the cost of a pipeline or injection equipment, it is common to include a nominal adder to account for the additional cost of transportation and storage (T&S). The 2007 DOE/NETL report did not include a T&S adder; however, this study examined the effect that a nominal $10/tonne of CO_2 T&S adder would have on the LCOE. The CO_2 T&S LCOE adder is calculated using the following equation:

$$CO_2\ T\&S = (CO_{2\,Cp\ captured})(CO_2\ T\&S\ cost)$$

where:

CO_2 T&S = cost of CO_2 transportation and storage adder for LCOE ($/MWh)
CO_2 = carbon dioxide (tonnes/MWh at plant capacity factor)
$_{Cp}$ = capture plant
CO_2 T&S cost = nominal cost adder for transportation and storage ($/tonne)

Exhibit 3 once again compares the LCOE and CO_2 mitigation costs of the MTR case and the MEA cases, this time with the additional CO_2 T&S cost. This affects both the LCOE and the CO_2 mitigation cost.

Exhibit 3
LCOE and CO_2 Mitigation Cost – Added CO_2 T&S Cost

	MTR	**MEA-1**	**MEA-1a**
Capital	32.2	25.7	25.7
Fixed O&M	1.7	0.9	0.9
Variable O&M	2.9	7.4	7.4
Coal	0.2	0.0	0.0
Natural Gas	0.0	0.2	0.2
Make-Up Power	20.2	21.4	18.7
CO_2 T&S	8.2	8.2	8.2
Total LCOE ($/MWh)	**65.3**	**63.8**	**61.2**
CO_2 Mitigation Cost ($/tonne)	**80.5**	**78.4**	**75.2**

COE and CO_2 Mitigation Cost – Added Make-up Power Emissions

Because of the reduced net power output of a plant with CO_2 capture, additional electricity is required from another source to match the original plant output. For this study, make-up power costs are included in the LCOE calculation and, as a result, the power output used for LCOE calculations is equivalent to the output of the base case plant. However, unless the make-up power comes from an emission-free source, like wind or solar, it will have CO_2 emissions associated with it. As mentioned in the Cost Estimate Basis, the make-up power cost assumed for this study was 7.1 ¢/kWh, which was assumed to be equivalent to a new subcritical pulverized coal plant. Therefore, the total electricity associated with the retrofit plant operation includes the CO_2 emissions of this make-up power. A third analysis of the CO_2 mitigation cost was conducted to include the CO_2 emissions of the make-up power as part of the CO_2 emitted by the retrofit plant. Exhibit 4 shows both the LCOE and CO_2 mitigation costs when these additional emissions are included. Appendix A shows the equations used for this calculation.

Exhibit 4
LCOE and CO_2 Mitigation Cost – Added Make-up Power Emissions

	MTR	MEA-1	MEA-1a
Capital	32.2	25.7	25.7
Fixed O&M	1.7	0.9	0.9
Variable O&M	2.9	7.4	7.4
Coal	0.2	0.0	0.0
Natural Gas	0.0	0.2	0.2
Make-Up Power	20.2	21.4	18.7
CO_2 T&S	8.2	8.2	8.2
Total LCOE ($/MWh)	**65.3**	**63.8**	**61.2**
CO_2 Mitigation Cost ($/tonne)	**117.8**	**117.8**	**106.4**

LCOE and CO_2 Mitigation Cost – Comparison of Approaches

Exhibit 5 compares the LCOE result sfor the three approaches described above. Including the CO_2 T&S cost in the LCOE calcualtions increases the LCOE, but including the emissions for the make-up power does not affect the LCOE results since there is no cost allowance for CO_2 emissions. Exhibit 6 compares the CO_2 mitigation cost results for the three approaches, in addition to showing the plant emissions. The increased LCOE when CO_2 T&S costs are included results in a similar increase in CO_2 mitigation costs. However, the much bigger effect is when the CO_2 emissions associated with the make-up power is included, which significantly decreases the emssions savings of the retrofit and in turn increases the mitigation costs. Because the MEA cases require more make-up power than the MTR case, the cost increase for including make-up power emissions is higher.

Exhibit 5
Comparison of LCOE Results

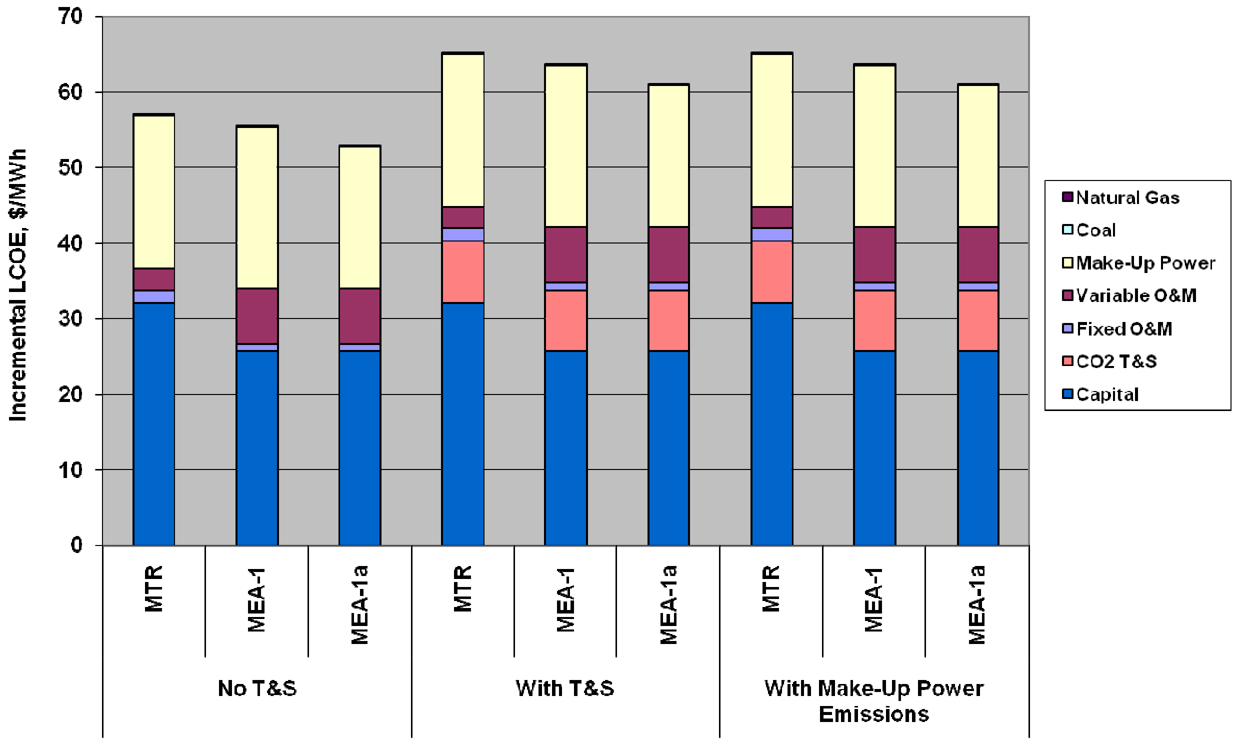
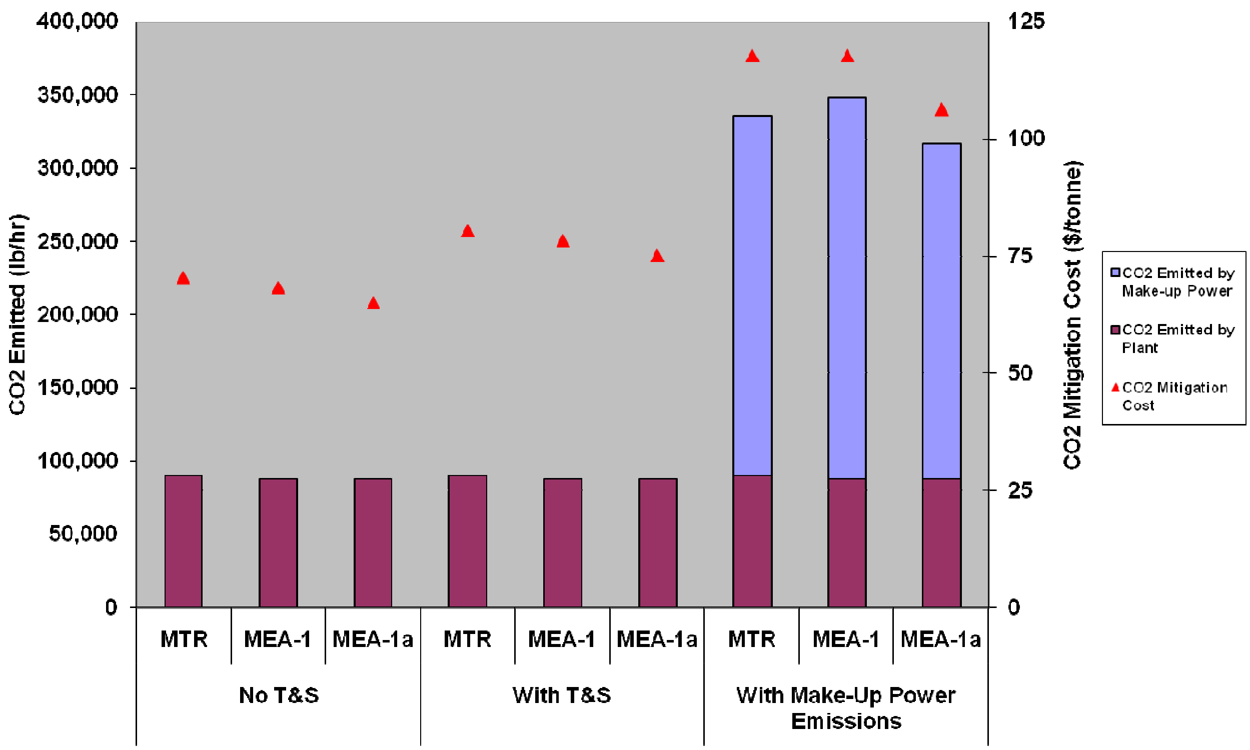


Exhibit 6
Comparison of Plant Emissions and CO_2 Mitigation Cost



LCOE and CO_2 Mitigation Cost – Sensitivity Cases

The LCOE and CO_2 mitigation costs were also calculated for the sensitivity cases evaluated. Exhibit 7 shows the LCOE results for the seven sensitivity cases, as well as the base MTR case for comparison. These cases were calculated using the base economic assumptions without accounting for the CO_2 T&S cost. Exhibit 8 shows the CO_2 mitigation cost results for the same seven cases plus the MTR base case. Again, the calculations were made with the base economic assumptions, not accounting for CO_2 T&S or the emissions associated with the make-up power. Because the coal flow rate for the different sensitivity cases does not change, the CO_2 emissions on a lb/hr basis does not change among the cases and, therefore, is not shown. However, the emissions on a lb/MWh basis would change due to the changes in plant efficiency and net plant output for some of the sensitivity cases.

The cases in the charts are abbreviated as followed:

MTR = MTR base case
High Turbo = High end of turbo machinery cost sensitivity (+20%)
Low Turbo = Low end of turbo machinery cost sensitivity (-20%)
High Membrane = High end of membrane cost sensitivity (+20%)

Low Membrane = Low end of membrane cost sensitivity (-20%)
Adv Membrane = Advanced membrane sensitivity
CO2 Purity = Relaxed CO_2 product specification sensitivity
Red Aux Load = Reduced auxiliary load sensitivity

Exhibit 7
LCOE for MTR Sensitivity Cases

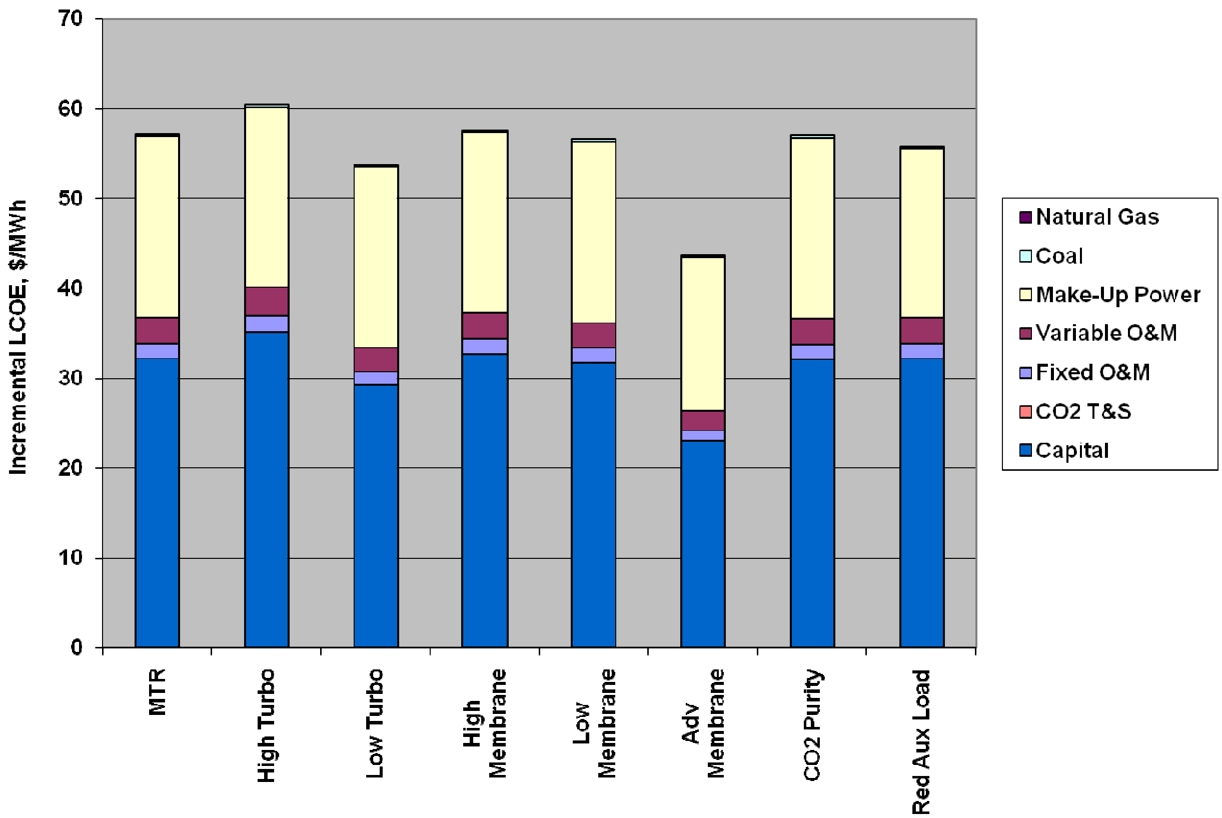
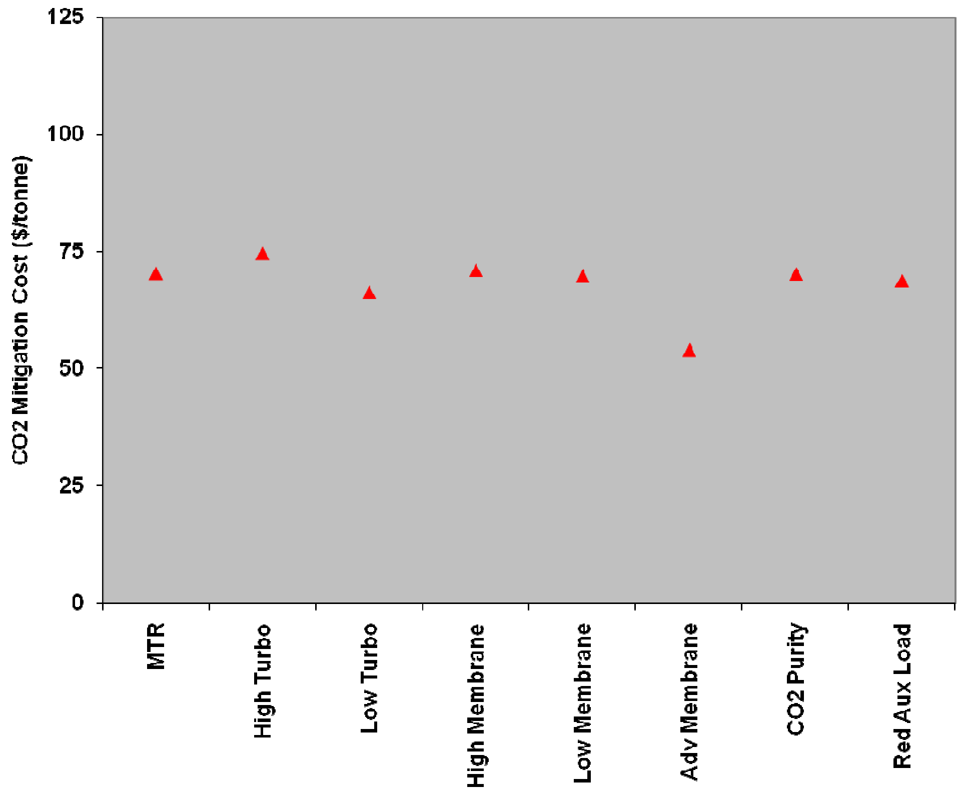


Exhibit 8
CO2 Mitigation Cost for MTR Sensitivity Cases



Appendix A

The annual make-up power cost is calculated as the product of the levelized cost of make-up power and the difference is annual output over the course of the year or:

$$Annual\ Makeup\ Power = [LCOE_{Makeup\ Power}][(CF_{ref})(MWh_{ref}) - (CF_{Cp})(MWh_{Cp})]$$

where:

Annual Makeup Power = annual cost of make-up power

$LCOE_{Makeup\ Power}$ = levelized cost of make-up power

CF = capacity factor

MWh = annual net megawatt-hours of power generated at 100% capacity factor (product of plant capacity in megawatts and 8,760 hours per year)

$_{Ref}$ = reference plant

$_{Cp}$ = capture plant

The annual make-up power cost is included in the variable O&M under the "Other" category as "Supplemental Electricity (for consumption)"

Final Report: March 2011

WorleyParsons Report No. EPRI-MTR 1-LI-011-0001-RB

WorleyParsons Job No. 108008-01604

Evaluation of MTR's CO2 Membrane Process for Capturing CO2 from Power Plant Flue Gas

Prepared for:

ELECTRIC POWER RESEARCH INSTITUTE

and

Membrane Technology & Research

Prepared by:

WorleyParsons resources & energy

WorleyParsons Group Inc.

March 25, 2011

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PROJECT: EVALUATION OF MTR'S CO2 MEMBRANE PROCESS FOR CAPTURING CO2 FROM POWER PLANT FLUE GAS							
REV	DESCRIPTION	ORIG	REVIEW	APPROVAL	DATE	CLIENT APPROVAL	DATE
A	Issued for Comment	David B. Stauffer D Stauffer	On file V Vaysman	David B. Stauffer D Stauffer, PM	31-Jan-11	email A. Bhown	23-Mar-11
Final	Issued as Final	David B. Stauffer D Stauffer	On file V Vaysman	David B. Stauffer D Stauffer, PM	25-Mar-11	N/A	

Table of Contents

List of Exhibits

Prepared by WorleyParsons Group
under contract to Electric Power Research Institute (EPRI)
in support of a Membrane Technology Research Project (MTR),
co-funded by the US Department of Energy (USDOE):

Principal Investigators for WorleyParsons include:

David Stauffer
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DOE Contract Number DE-NT0005312

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Mr. Ramadoss Thiagarajan, Senior Process Engineer

Acronyms and Abbreviations

AACE	Association for the Advancement of Cost Engineering
acfm	Actual cubic feet per minute
amsl	above mean sea level
AR	As received
atm	Atmosphere (14.696 psi)
BACT	Best available control technology
BEC	Bare erected cost
BFD	Block flow diagram
Btu	British thermal unit
Btu/hr	British thermal units per hour
Btu/kWh	British thermal units per kilowatt hour
Btu/lb	British thermal units per pound
CCOFA	Close coupled over-fired air
cf	Cubic feet
CF	Capacity factor
CFM	Cubic feet per minute
CO	Carbon monoxide
CO_2	Carbon dioxide
CRH	Cold reheat
CS	Carbon steel
dB	Decibel
DB	Dry basis
DCC	Direct contact cooler
DOE	Department of Energy
EPA	Environmental Protection Agency
EPC	Engineer/Procure/Construct
EPRI	Electric Power Research Institute
EPCM	Engineering/Procurement/ Construction Management
ESP	Electrostatic precipitator
FD	Forced draft
FG	Flue gas
FGD	Flue gas desulfurization
FOAK	First of a kind
ft	Foot, Feet
FW	Feedwater
FO&M	Fixed operations and maintenance
gal	Gallon
gal/MWh	Gallons per megawatt hour
GJ	Gigajoule
GHG	Greenhouse gas
GPM	Gallons per minute
h, hr	Hour
H_2O	Water
HHV	Higher heating value
Horiz.	Horizontal
hp	Horsepower
HP	High pressure
HRH	Hot reheat
Hz	Hertz
ID	Induced draft
in. H_2O	Inch water
in. Hga	Inch mercury (absolute pressure)
in. W.C.	Inch water column
IP	Intermediate pressure
KO	Knockout
kV	Kilovolt
kW, kWe	Kilowatt electric
kWh	Kilowatt-hour
kWt	Kilowatt thermal
lb	Pound
lbm	Pounds, mass
lb/hr	Pounds per hour
lb/MMBtu	Pounds per million British thermal units
lbmol	Pound mole
lbmol/hr	Pound moles per hour
lb/MWh	Pounds per megawatt hour
lb/TBtu	Pounds per trillion British thermal units
LP	Low pressure
LV	Low voltage
MCR	Maximum continuous rate
MEA	Monoethanolamine (CO_2 scrubber solvent)
MM	Million
MMBtu	Million British thermal units
mole%	Mole percent (percent by mole)
MTR	Membrane Technology Research
MUPC	Make-up power cost
MVA	Mega volt-amps
MW,MWe	Megawatt electric
MWh	Megawatt-hour
MWt	Megawatt thermal
N_2	Nitrogen
N_2O	Nitrous oxide
N/A	Not applicable

NAAQS	National Ambient Air Quality Standards
NETL	National Energy Technology Laboratory
NOx	Oxides of nitrogen
NRE	Non-recurring engineering (costs)
NSPS	New Source Performance Standards
NSR	New Source Review
O_2	Oxygen
O&M	Operation and maintenance
OD	Outside diameter
OP/VWO	Over pressure/valve wide open
PA	Primary air
PC	Pulverized coal
PF	Power factor
PFD	Process flow diagram
PM	Particulate matter
PM_{10}	Particulate matter measuring 10 µm (micrometers) or less
$PM_{2.5}$	Particulate matter measuring 2.5 µm (micrometers) or less
ppm	Parts per million
ppmv	Parts per million volume
ppmvd	Parts per million volume, dry
PSD	Prevention of significant deterioration
psi	Pounds per square inch
psia	Pound per square inch absolute
psid	Pound per square inch differential
psig	Pound per square inch gage
Qty	Quantity
RH	Reheater
Rpm	Revolutions per minute
SA	Secondary air
scf	Standard cubic feet
scfd	Standard cubic feet per day
scfh	Standard cubic feet per hour
scfm	Standard cubic feet per minute
scmh	Standard cubic meter per hour
SG	Specific gravity
SO_2	Sulfur dioxide
SOA	State-of-art
SOFA	Separated over-fire air
SOx	Oxides of sulfur
SS	Stainless steel
st	Short ton
STG	Steam turbine generator
TEG	Tri-Ethylene Glycol
TIC	Total Investment Cost
tonne	Metric ton (1,000 kg)
TPC	Total plant cost
TPD	Ton per day
TPH	Tons per hour
TPI	Total plant investment
U.S.	United States
V	Voltage
VO&M	Variable operations and maintenance
Vert.	Vertical
V-L	Vapor liquid portion of stream (excluding solids)
vol%	Volume percent
VWO	Valves wide open
WB	Wet bulb
wg	Water gauge
wt%	Weight percent
$/MMBtu	Dollars per million British thermal units
$/kW	Dollars per kilowatt
°C	Degrees Celsius
°F	Degrees Fahrenheit

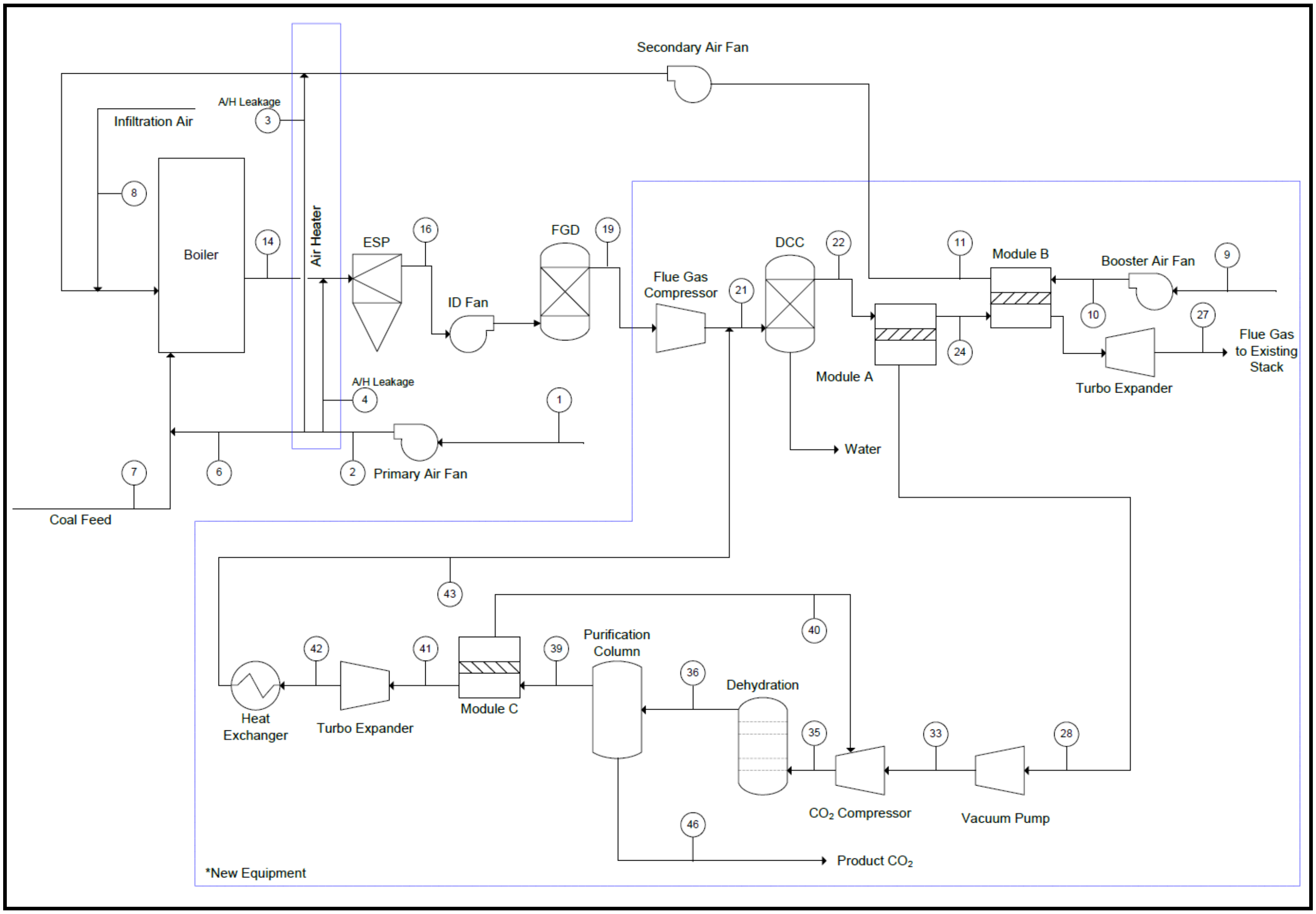
EXECUTIVE SUMMARY

The objective of this evaluation by WorleyParsons is to provide technical and cost input to EPRI for the economic evaluation of MTR's CO_2 membrane capture system in a retrofit application to a representative pulverized-coal (PC) plant. AEP's Conesville Unit 5 was chosen as the subject PC unit for the evaluation. This membrane retrofit application was designed to capture 90% of the CO_2 generated, paralleling an earlier study that evaluated CO_2 capture via an amine-based capture system on the Conesville Unit 5, thus facilitating the comparison of the membrane and amine capture systems.

Conesville Unit 5 is a 2400 psig class Rankine cycle with nominal steam turbine throttle characteristics of 2400 psig/ 1000°F /1000°F. The base case scenario for Conesville Unit 5 is based on 5% overpressure and valves wide open (VWO) operation, with a gross generation of 463,478 kW, net generation of 433,778 kW, and an auxiliary load of 29,700 kW. The net plant heat rate is 9,749 Btu/kWh, with a corresponding net plant efficiency of 35.01% (HHV). The corresponding values for the membrane based CO_2 capture process are: gross generation of 486,896 kW, net generation of 310,542 kW, auxiliary load of 176,354 kW, net plant heat rate of 13,762 Btu/kWh, and net plant heat rate of 24.80%.

MTR's CO_2 membrane capture system centers around the addition of CO_2 permeable membranes into the flue gas stream downstream of the existing FGD. A simplified block flow diagram is presented in Exhibit ES-1. A flue gas compressor is added to facilitate operation of the membranes at 2 bar to minimize the required membrane area and to improve the CO_2 separation. CO_2 that is separated by the first membrane modules, which are cross flow modules, is sent for compression and additional purification. The CO_2 specification of less than 100 ppm O_2 in the CO_2 product requires the introduction of a purification column. A small stream rejected by the compression / purification system is recycled back to the CO_2 separation system. Flue gas retentate leaving the first CO_2 module passes through the counter flow modules where fresh air is used to sweep the permeate side of the membrane. The sweep air / permeate leaving the counter flow membrane is integrated into the secondary air system of the existing PC steam generator.

Exhibit ES-1
CO_2 Membrane Process Block Flow Diagram



Note: Module A consists of cross flow membranes. Module B consists of counter flow sweep membranes.

Compared to the base case secondary air, the sweep air fed to the secondary air system is vitiated of oxygen (circa 17% O_2), has an increased CO_2 level (circa 9% CO_2) and overall has increased mass and volumetric flows of about 19% and 25% respectively. These changes to the secondary air lower the boiler efficiency by approximately 0.8%. The largest performance effect of the retrofit on Conesville Unit 5 is the significantly increased auxiliary load associated with the required turbo-machinery. Additional discussion of the effects of the retrofit on the plant are presented in Section 4.

A summary of the cases compared in this study is presented in Exhibit ES-2. The cases highlighted with a light green background were developed in a standalone study addressing the amine based retrofit and discussed further in Section 1.2.1.

Exhibit ES-2 Evaluation Matrix

Case	Description	CO_2 Capture/ Compression	Cost	Notes
Base-0	Do Nothing Case (Existing Facility)	None	NA	Existing Conesville Unit No. 5.
MTR-1	MTR CO_2 Membrane Retrofit	90% capture/ 2015 psia	Dec 2009 $	Retrofit of Conesville Unit No. 5 [Focus of this Evaluation.]
MEA-1	MEA Retrofit Retrofit (SOA 2006)	90% capture/ 2015 psia	Escalate to Dec 2009 $	Solvent regeneration energy of 1550 Btu/lbm-CO_2. (Note a)
MEA-1a	MEA Retrofit Retrofit (Advanced)	90% capture/ 2015 psia	Cost presumed to be equivalent to MEA-1	Solvent regeneration energy of 1200 Btu/lbm-CO_2. (Note a)

Note a. The MEA-1 and -1a Retrofit cases are known as "Case 1" and "Case 1a" within Reference [1].

The technical and cost information developed for the evaluation of the CO_2 membrane retrofit are based on a conceptual level of detail.

PERFORMANCE SUMMARY

A performance summary for the four cases listed above is presented in Exhibit ES-3. The MTR CO_2 membrane capture system is fully capable of capturing the targeted 90% of the CO_2 generated. The impact to the generation unit is such that the gross generation increased by approximately 23,418 kW to 486,896 kW, while the auxiliary load increased by approximately 146,654 kW to 176,354 kW, yielding a net generation of 310,542 kW which is down 123,236 kW from the base case value of 433,778 kW. Compared to the MEA-1 case net generation of 303,310 kW, the MTR-1 retrofit case has a 7,232 kW advantage over MEA-1. Compared to the MEA-1A case net generation of 319,280 kW, the MTR-1 retrofit case has a 8,738 kW disadvantage. The reader is reminded that the MEA-1 case was the state-of-the-art MEA application at the time of the reference study, and that the MEA-1A was the advanced MEA application, not based on specific technology advances, nor was its cost estimated. Additional discussion about the MEA study can be found in Section 1.2.1.

Exhibit ES-3
Summary of Technical Performance for Retrofitting Conesville Unit 5

Parameter	Units	Case-0	MTR-1	MEA-1	MEA-1A
		Base Case	SOA 2010	SOA 2006	Advanced
Boiler Parameters					
Boiler Efficiency	percent	88.13	87.33	88.13	88.13
Coal Heat Input (HHV)	MMBtu/hr	4,229	4,274	4,229	4,229
CO2 Removal System Steam & Related Parameters					
Solvent Regeneration Energy	Btu/lbm-CO2	-	NA	1,550	1,200
CO2 Removal System Steam Extraction Flow	lbm/hr	---	5,696	1,210,043	975,152
Natural Gas Heat Input	MMBtu/hr	-	-	13	13
Generation & Auxiliary Load					
Existing Steam Turbine Generator Output	kW	463,478	463,044	342,693	367,859
CO2 Removal System Turbine Generator Output	kW	-	23,852	45,321	36,083
Total Turbine Generator Output	kW	463,478	486,896	388,014	403,942
Auxilliary Power: Existing Plant	kW	29,700	33,430	29,765	29,817
Auxilliary Power: CO2 Removal System	kW	-	142,924	54,939	54,845
Net Plant Output	**kW**	**433,778**	**310,542**	**303,310**	**319,280**
Plant Performance Parameters					
Net Plant Heat Rate (HHV)	Btu/kWh	9,749	13,762	13,985	13,285
Net Plant Efficiency (HHV)	%	35.01%	24.80%	24.41%	25.69%
Energy Penalty, (percentage points of NP Eff.)	%	Base	10.21%	10.60%	9.32%
Plant CO2 Emissions					
Carbon Dioxide Produced	lbm/hr	866,102	872,184	867,595	867,595
Carbon Dioxide Recovered	lbm/hr	-	782,177	779,775	779,775
Carbon Dioxide Emissions	lbm/hr	866,102	90,007	87,820	87,820
Carbon Dioxide Recovered (% of Produced)	%	0.00%	89.68%	89.88%	89.88%
Specific Carbon Dioxide Emissions	lbm/kWh	1.997	0.290	0.290	0.275
Normalized Specific CO2 Emissions (Relative to Base Case)	fraction	1.000	0.145	0.145	0.138
Avoided Carbon Dioxide Emissions (as compared to Base)	lbm/kWh	---	1.707	1.707	1.722

Note: Reference [1] page 152, Table 5-2 is the source of values for Base Case, MEA-1, and MEA-1A.

COST ESTIMATING SUMMARY

The capital cost estimates developed herein have an accuracy level of ±30%, consistent with the conceptual level of the study. The results of the capital and O&M cost estimation effort are represented in Exhibit ES-4. The amine reference study [1] developed costs in July 2006 USD. As part of this study, WorleyParsons escalated the MEA costs to December 2009 to facilitate the comparison with the membrane retrofit cost developed herein. Nevertheless, care should be taken in comparing the MTR-1 case costs with the MEA-1 costs since two different organizations were responsible for the development of the costs. When costs are developed within a given study, there is a higher degree of consistency in the cost model development than between costs developed in different studies. Said differently, the relative accuracy between the cost estimates is greatest when developed by the same cost estimating model developed at the same moment in time. This topic is discussed further in Section 6.1.

Per Exhibit ES-4 the Total Investment Cost (TIC) of the MTR-1 case is 25% higher than the MEA-1 case on a dollars basis, or 22% higher on a dollar per kW basis. Again care should be taken comparing a 20-25% difference for two values that are approximately ±30% in accuracy and developed for different studies with different cost models.

Exhibit ES-4 Capital & O&M Cost Summary (Dec 2009 dollars)

Parameter	Units	Case-0	MTR-1	MEA-1	MEA-1
			Dec 2009 $	Jul 2006 $	Dec 2009 $
Total Investment Cost	$1,000	Base	594,484	400,094	474,940
Total Investment Cost	$/kW, net	Base	1,914	1,319	1,566
Fixed O&M Costs	$1000/yr	Base	4,681	2,494	2,647
Variable O&M Costs	$1000/yr	Base	7,963	17,645	20,631
Levelized, Makeup Power Cost	$1000/yr	Base	65,151	62,194	68,996
CO2 byproduct Revenue	$1000/yr	Base	-	-	-
Feedstock (Natural Gas) O&M Cost	$1000/yr	Base	-	653	575

Note: Costs for MEA-1 (Y2006 USD) are based on Reference [1], p ES-3.

A high-level sensitivity study for an advanced membrane system shows the potential of saving approximately $168 million while producing an additional 18 MW of net power is documented in Section 5.4.2. MTR may wish to pursue this option further in the future.

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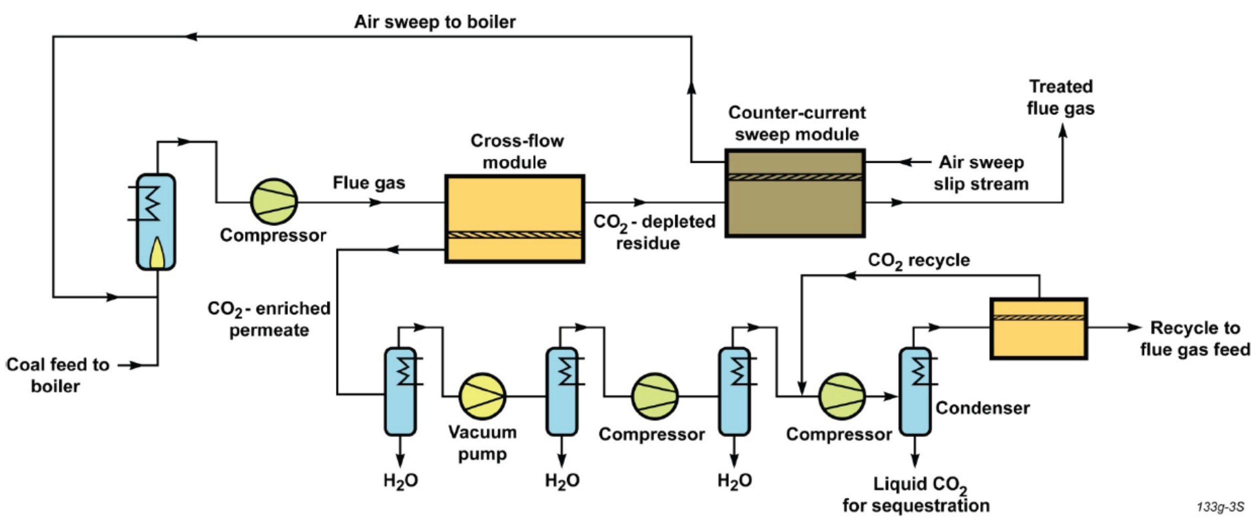
1. INTRODUCTION

The overall objective of this evaluation by WorleyParsons is to provide technical and cost input to EPRI for the economic evaluation of MTR's CO_2 membrane capture system, based on MTR's field and lab data.

The objective of the MTR CO_2 membrane capture system is to capture 90% of the CO_2 from the flue gas of a traditional pulverized coal (PC) plant while minimizing the auxiliary load through implementation of the MTR cross flow and counter flow modules with a vacuum and sweep air respectively. A high level depiction of this process is presented in Exhibit 1-1.

Exhibit 1-1
MTR CO_2 Membrane Application to a Pulverized Coal Plant – A High Level BFD



The heart of the CO_2 membrane process is the cross-flow module and the counter-current sweep module, both of which operate on a partial pressure driving force. The cross flow membrane achieves the CO_2 partial pressure driving force through a pressure gradient achieved by balancing feed compression and a permeate vacuum. Although vacuum pumps may be less efficient than gas compressors, the vacuum pump contributes to the overall process efficiency since the permeate flow is a small fraction of the feed flow.

The counter-current sweep module achieves its CO_2 partial pressure driving force through the use of an air sweep stream. The advantage of using the air sweep stream is that the process air can be utilized in the PC boiler, and that the entrained CO_2 is recycled back into the process, thus increasing the overall capture rate[a].

[a] Where less than 90% CO_2 capture is required, a membrane process that eliminates the counter-current sweep module could be entertained, thus eliminating the retrofit impact on the steam generator and secondary air system. If such a capture system were desired, this has the potential to be a significant sweet spot for the membrane system.

The other unit operations seen in the Exhibit 1-1 work together to compress and purify the captured CO_2.

A summary of the cases to be compared in this study is presented in Exhibit 1-2. The cases highlighted with a light green back ground were developed in a standalone study discussed further in Section 1.2.1.

Exhibit 1-2 Evaluation Matrix

Case	Description	CO_2 Capture/ Compression	Cost	Notes
Base-0	Do Nothing Case (Existing Facility)	None	NA	Existing Conesville Unit No. 5
MTR-1	MTR CO_2 Membrane Retrofit	90% capture/ 2015 psia	Dec 2009 $	Retrofit of Conesville Unit No. 5 [Focus of this Evaluation.]
MEA-1	MEA Retrofit Retrofit (SOA 2006)	90% capture/ 2015 psia	Escalate to Dec 2009 $	Solvent regeneration energy of 1550 Btu/lbm-CO_2. (Note a)
MEA-1a	MEA Retrofit Retrofit (Advanced)	90% capture/ 2015 psia	Cost presumed to be equivalent to MEA-1	Solvent regeneration energy of 1200 Btu/lbm-CO_2. (Note a)

Note a. The MEA-1 and -1a Retrofit cases are known as "Case 1" and "Case 1a" within Reference [1].

1.1 PROJECT OBJECTIVE

The objectives for MTR CO_2 membrane retrofit project include the following.

- The CO_2 membrane capture application will be based on retrofitting American Electric Power (AEP) Conesville Unit No. 5 as documented by the DOE/NETL 401/110907 Report, entitled "Carbon Dioxide Capture from Existing Coal-Fired Power Plants, November 2007 [1, 2].
- The carbon capture basis will be 90%, based on the produced carbon dioxide. (Sec 2.7, CO_2)

This report will focus on the technical aspects of the retrofit and the system integration, and development of capital and operating cost. EPRI will take the costs and develop an economic analysis and comparison to the MEA retrofit options presented in Exhibit 1-1.

1.2 STUDY BACKGROUND / APPROACH

This section will highlight the relevant background of the reference study [1] and discuss at a high-level, the modeling approach utilized in this analysis.

1.2.1 Comparison Amine Study

If the US power industry is required to take action to reduce the Greenhouse Gas (GHG) emissions from power generating facilities, it is likely that existing plants would need to be retrofitted. Presently the state-of-the-art CO_2 emission control technology for retrofitting existing plants is the amine-based CO_2 scrubbing technology. It is with this technology that the MTR CO_2 membrane technology must directly compete.

A recent NETL study [1] looked at the impact of retrofitting an existing PC-fired power plant with an amine-based CO_2 scrubbing system. This study evaluated the technical, cost and economic impacts of removing CO_2 from the post combustion flue gas with an advanced amine scrubbing system at the Conesville Unit 5 Plant. The study evaluated various levels of CO_2 capture (0%, 30%, 50%, 70% and 90% capture), as well as a sensitivity study showing the effect of possible reductions in the solvent regeneration energy (1550 and 1200 Btu/lb-CO_2) for the 90% capture case. The 1550 Btu/lb-CO_2 case represents the state-of-the-art MEA technology at the time of the study (circa 2006). The 1200 Btu/lb-CO_2 level represents a near future value that may be achievable with improved solvent and other unspecified technological improvements. Additional discussion regarding the MEA evaluation is presented in Section 2.8.

In light of the interest in comparing the state-of-the-art MEA retrofit to the CO_2 membrane system retrofit, it was decreed early in the present membrane study to utilize the same basis as the reference MEA study. Thus, the present study will be based on retrofitting Conesville Unit 5, capturing 90% of the generated CO_2, and performing the retrofit evaluation on the same equipment limitations/basis as documented in reference [1].

A comparison of the MTR membrane retrofit to the MEA retrofit results is made in Section 6.

1.2.2 Conesville Unit 5

Conesville Unit 5 is a nominal 450 MW reheat, subcritical, pulverized-coal (PC) fired steam plant operated by AEP of Columbus Ohio. Unit 5 is one of six coal fired PC steam plants located on the Conesville site with a total generating capacity of approximately 2,080 MW. The Unit 5 steam generator is a reheat unit with controlled circulation, a single furnace cell employing corner firing and tilting tangential burners. The utilized fuel is a bituminous coal from Ohio. The flue gas leaving the steam generator is cleaned by an Electrostatic Precipitator (ESP) and a lime-based flue gas desulfurization (FGD) system before being discharged to the atmosphere.

The steam turbine generator employs nominal steam conditions of 2400 psig/ 1000°F / 1000°F, exhausts to a condenser back pressure of approximately 2.5 inches Hga, and has been designed for 105% overpressure operation. The unit heat rejection is accomplished by two five-cell mechanical draft evaporative cooling towers.

The Conesville Unit 5 is representative in many ways of a large number of pulverized coal-fired units in use today throughout the US. As such, it is an excellent unit for the subject of this CO_2 membrane retrofit application.

1.2.3 Modeling Approach

A critical input for determining the impact of the CO_2 membrane retrofit on Conesville Unit 5 is the development of the heat and mass balance and corresponding performance estimate. To this end, several different specialized computer modeling software programs were employed, each with its own niche in the overall analysis. The modeling software is listed below, followed by a brief description of how it was utilized within the analysis.

- **MTR proprietary membrane software** – MTR provided the performance for each of the three different membranes utilized in the evaluation. The software accounts for the membrane operating conditions, permeability, inlet composition, pressure ratios or sweep air flow rate, and geometry. The information provided by MTR was utilized by WorleyParsons in the supplemental analyses.

- **ASPEN** – WorleyParsons utilized the ASPEN software to evaluate the impact of the membrane retrofit to the boiler, air and flue gas gaseous streams. The ASPEN analysis is complicated by the presence of two recycle streams: the sweep air from the cross flow module and the CO_2 purification system recycle stream. The presence of the recycle streams required that WorleyParsons and MTR iterate between their modeling software to ensure sufficient convergence of the results. Since the majority of the membrane retrofit impact is to the unit's gas side, the ASPEN analysis represents the heart of the overall analysis.
- **Boiler Performance Model (BPM)** - WorleyParsons utilized the BPM model to address the effect of the increased CO_2 and N_2 flowing through the boiler as a result of the membrane's sweep vitiated air feeding the secondary air fans in lieu of fresh air. This simulation was not originally envisioned in the project scope, but was performed after the OEM for the steam generator declined to participate in the study. The BPM software was utilized primarily to determine an approximate impact to the boiler efficiency as well as to gain a preliminary understanding into how sufficient the existing temperature control schemes are in view of the redistribution of heat absorption within the boiler resulting from the membrane retrofit.
- **GATE**: WorleyParsons utilized the GATE software to address the impacts to the steam turbine cycle resulting from integration of the membrane system. Initially the GATE program was utilized in evaluation of the process heat integration concepts that were ultimately rejected as not being justifiable. In the end, the only change to the steam turbine cycle was accounting for a new steam extraction required by the CO_2 drying process.

The above models were exercised many times over the course of the study as the process assumptions were refined, and as the process models were interfaced and converged. In addition, preliminary analyses were developed to help address issues related to the optimization of the overall process. For example, work was performed which identified that operating the membranes at 2 bar consumed less parasitic power and required significantly fewer membrane modules (i.e., less space and cost) than if the membranes were operated at 1.2 bar.

1.3 REPORT STRUCTURE

This report is structured according to the following major section outline:

- Executive Summary
- Section 1: Introduction
- Section 2: Evaluation Basis
- Section 3: Conesville Unit 5 Description
- Section 4: CO_2 Membrane Retrofit Description & Performance
- Section 5: Cost Analysis of Retrofit
- Section 6: Results

2. EVALUATION BASIS

This section contains a summary of essential technical and functional requirements that were used as a basis in establishing the conceptual design for the CO_2 membrane retrofit case evaluated in this report. Further design criteria details are provided in the Design Basis Document for the MTR CO_2 Membrane Retrofit of the AEP Conesville Unit 5 found in Appendix D. [3]

2.1 SITE DESCRIPTION

The ambient conditions for the project are assumed according to the following bases:

- Equipment design is based on site conditions for Conesville, OH; and
- Process modeling work is based on the American Boiler Manufacturers Association (ABMA) standard conditions.

The heat and mass balances are evaluated at the ABMA standard conditions, while equipment (e.g., vacuum pumps, fans, compressors, etc.) is sized for the indicated site conditions. This way, the performance estimate will facilitate a comparison of this study with the earlier completed Conesville Unit 5 carbon dioxide capture retrofit study [1]. Design and cost analyses are based on site specific conditions.

Conesville station site ambient conditions assumed in this study are based on reference [1], and presented in Exhibit 2-1.

Exhibit 2-1
Site Characteristics

Characteristic	Units	Value
Location		Conesville, Coshocton County, Ohio
Elevation, amsl	ft	744
Barometric Pressure	psia	14.31
Design Ambient Temperature, Wet Bulb	°F	75
Mean Coincident Dry Bulb Temperature (Note)	°F	85
Dry Bulb Temperature Extremes		
Maximum	°F	92
Minimum	°F	-1
Average Cooling Tower Water Temperature	°F	80

Note: 1% ASHRAE for Columbus, OH airport, [4]

Plant performance and heat and mass balances in this study will be referenced to the ABMA standard conditions [1] as presented in Exhibit 2-2.

Exhibit 2-2
ABMA Standard Conditions

Characteristic	Units	Value
Barometric Pressure	psia	14.696
Ambient Temperature, Dry Bulb	°F	80
Relative Humidity	%	60

The dry ambient air composition [5] utilized in this analysis is as presented in Exhibit 2-3.

Exhibit 2-3
Dry Ambient Air Quality

Constituent	Chemical Formula	Mole %, dry
Nitrogen	N_2	78.08%
Oxygen	O_2	20.95%
Argon	Ar	0.93%
Carbon Dioxide	CO_2	0.03%
	Total	100.00%

2.2 COAL ANALYSIS

An analysis of as-received mid-western bituminous coal that is currently being utilized by the Conesville Unit 5 boiler is presented in Exhibit 2-4. [1]

Exhibit 2-4
Design Coal

Parameter	Units	Value
Proximate Analysis		
Moisture	wt %	10.1
Ash	wt %	11.3
Volatile Matter	wt %	32.7
Fixed Carbon	wt %	45.9
Total	wt %	100.0
Ultimate Analysis		
Moisture	wt %	10.1
Ash	wt %	11.3
H	wt %	4.3
C	wt %	63.2
S	wt %	2.7
N	wt %	1.3
O	wt %	7.1
Total	wt %	**100.0**
Higher Heating Value	Btu/lb	11,293

2.3 SORBENT ANALYSIS

An analysis of lime that is currently being utilized by the Conesville Unit 5 Flue Gas Desulfurization (FGD) system is presented in Exhibit 2-5 [1].

Exhibit 2-5
Limestone Analysis

Constituent	Units	Value
CaO	wt%	90%
MgO	wt%	5%
Inerts	wt%	5%

2.4 ENVIRONMENTAL REQUIREMENTS

The Conesville Generating Station is located near Conesville, Coshocton County, in the state of Ohio. Coshocton County is located in the Zanesville-Cambridge Intrastate Air Quality Control Region (AQCR-183) and is currently designated by EPA (40 CFR 81.336) as an attainment area for all criteria pollutants (however, a portion of Coshocton County [Franklin Township] is designated non-attainment for $PM_{2.5}$); that is, the air quality in Coshocton County meets, or exceeds National Ambient Air Quality Standards (NAAQS).

2.4.1 Emissions Design Criteria Summary

In general, units that undergo physical changes or changes in the method of operation will be subject to New Source Performance Standards (NSPS) or PSD/NSR regulations if there is an increase in the maximum hourly emissions or a significant net increase in annual emissions, respectively. However, the membrane retrofitted plant is expected to operate at or below current emission levels of the existing Conesville Unit 5. The incorporation of the CO_2 membrane technology is expected to eliminate the emissions of 90% of the carbon dioxide and is not expected to increase any individual emission. In fact the CO_2 membrane system is expected to reduce the plant weight based emissions of particulate matter (PM) , SO_2, and NOx.

Therefore the retrofit project is based on utilizing the existing Unit 5 control technologies at the Conesville Power Plant rather than installing "new" Best Available Control Technology (BACT) controls. While the emission levels will not play a role in determining retrofit strategy in this study, the emissions resulting from the Conesville CO_2 membrane retrofit project are estimated and compared to the existing Unit 5 emissions, as well as the reference Amine CO_2 capture study [1].

A description of the existing combustion equipment, the existing Title V Final Air Permit limits and a list of the air quality control equipment that has been installed on Unit 5 at the Conesville Station are shown in Exhibit 2-6.

Exhibit 2-6
Existing Unit 5 Equipment, Emission Limits, and Air Quality Control Technology

Parameter	Units	Value
Fuel		Bituminous Coal
Nominal Heat Input	MMBtu/hr	4,091 (Note)
Gross Electric Output	MWe	450 [1]
PM Permit Limit	lb/MMBtu	0.10 [6]
PM Control Equipment		ESP
Fuel Sulfur Content	%	2.4 - 3.2
SO_2 Permit Limit	lb/MMBtu	1.2 [7]
SO_2 Control Equipment		Wet FGD, Lime
NOx Permit Limit	lb/MMBtu	0.45 [7]
NOx Control Technology		SOFA

Note: The Title V Final Air Permit for Unit 5 Main Boiler at the Conesville Power Plant "…having a nominal capacity of 4,091 MMBtu/hr…"

2.4.2 Carbon Reduction Requirements

Carbon dioxide (CO_2) emission is not currently regulated. As such, the requirements for CO_2 capture and the CO_2 product specification are not definitively defined. Since one objective of this analysis is the comparison of the CO_2 membrane retrofit to the MEA retrofit study [1], the CO_2 specifications for this project are based largely on the reference study. In general, the basis from that study is utilized except that the sulfur limit is presented herein as SO_2 in lieu of H_2S, and the moisture limit is based on NETL values instead of an average operating value for

Rectisol as was utilized in the reference study. Additional detail behind the development of the CO_2 specifications can be found in the Project Design Basis Document (Appendix D), section 2.7.

The implemented product CO_2 specification at the pipeline inlet at the plant boundary is presented in Exhibit 2-7. [8]

Exhibit 2-7
Product CO_2 Specification

Parameter	Units	Value	Notes
CO_2 Product End Use		EOR	
Pressure	psia	2,015	Supercritical fluid
CO_2, min	vol%	96%	Immaterial. Expect 99%
H_2O, max	vol%	0.015	i.e., 150 ppm. TEG dehydration capable of 100ppm Molecular sieve dehydration capable of <0.1 ppm.
N_2, max	vol%	0.6%	Immaterial. Expect 0.1%
O_2, max	ppmv	100	
SO_2, max	vol%	1%	Ref study [1] specified 1% H_2S. 3% would be relevant for geological sequestration.

The carbon reduction basis is a nominal 90 percent removal based on carbon input from the coal and excluding residual carbon that exits the boiler with the ash. An alternate way of describing this CO_2 capturing basis, is simply 90 percent removal of the produced carbon dioxide.

2.5 MTR CO_2 MEMBRANE REQUIREMENTS AND PERFORMANCE

The membrane requirements and performance are presented within this section.

The preliminary design requirements for the MTR CO_2 membrane requirements for Conesville are presented in Exhibit 2-8.

Exhibit 2-8
MTR CO_2 Membrane Inlet Requirements – Presumed for Conesville

Criteria	Limit/ Target	Note
PM	Not yet known	The particulate matter is the **greatest concern** of all of the anticipated contaminants, as it can lead to life ending fouling/ clogging. As such, the PM limit will be discussed in a dedicated subsection below.
SOx	NA	The membrane is robust with respect to SOx. Both SO_2 and SO_3 are polar and will permeate through the membrane.
NOx	NA	The membrane is robust with respect to NOx. No limit was specified.
O_2	NA	There is no limit on the O_2 in the flue gas stream.
T, Feed	<70^oC	The membrane is robust at temperatures below 70^oC. The preferred range is 10-50^oC for better membrane performance.
T, Superheat	NA	Flue gas can be saturated with water when fed into the first membrane step. Since the membrane is very permeable to water, the water content in the feed decreases rapidly. As a result, water condensation on the feed side of the membrane appears unlikely. The concern with liquid condensation is that it would block flow channels causing undesirable pressure drop.
P, Feed	**2.0 bar**	Analysis by MTR [9] indicates that the membrane area is reduced most appreciable by being between 2 and 3 bars, while the net power will be notably higher at 2 bar versus 3 bars. As such, the 2 bar pressure will form the basis of this analysis.
Flow rate	Design Flow	The retrofit will be designed to handle 100% of the flue gas flow. For lower capture rates, membrane bypass can be used. However, since 90% capture is the target, a bypass is not envisioned.
Heavy Metals	NA	MTR membrane is not adversely affected by heavy metals.

Reference: [10], unless noted otherwise within table.

2.5.1 Particulate Matter Membrane Requirement

Particulate matter is a potential membrane contaminant of much interest to the project members as deposition of the PM may lead to fouling and clogging of the membrane. Unfortunately there is great uncertainty regarding how much PM will deposit in the membrane versus simply pass through the membrane. High efficiency candle filters could be added, but they add a significant pressure drop, require substantial real estate and are costly. On top of that, the candle filters may not be necessary. Similarly a wet ESP could also be added, but space and cost would likely become issues.

The membrane flow path is measured on the order of a millimeter. Although no information is presently available, the particulate size leaving the ESP/FGD is postulated to be on the order of several microns to submicron. Since 1000 microns fits between a 1 mm flow path, it seems possible that much of the ash could be carried thorough the entire membrane. The validity of this hypothesis will be examined as part of the Cholla CO_2 Membrane demonstration project.

Therefore, this first phase of evaluating the membrane retrofit application will be based on the assumption that high efficiency candle filters downstream of the ESP/FGD are not needed. [11]

2.5.2 MTR CO_2 Membrane Performance Parameters

The CO_2 membrane performance data was extracted from simulation data provided by MTR and incorporated into the heat and mass balance developed for the project. The details of the membrane performance is confidential to MTR and will not be presented in this report. For this reason, some composition information has been redacted from the developed H&MB prior to presentation herein. General performance observations regarding the membrane are presented below.

The MTR CO_2 membrane is based on the Polaris membrane which allows polar molecules (e.g., H_2O, CO_2, SO_2, SO_3, H_2S, NO_2) to permeate. Although SO_3 is expected to permeate through the CO_2 membrane into the CO_2 product, being a hydrophilic molecule it would end up with the water removed by the CO_2 compression process.

Oxygen gas, O_2, a non polar molecule, will preferentially be rejected by the membrane, and depending upon the feed concentration, could comprise up to approximately 1.5% of the CO_2 product stream leaving the cross flow membrane.

The O_2 content of the sweep air leaving the counter-flow membrane module is approximately 17-18% O_2. That is, some O_2 in the incoming air leaks through to the flue gas side, thus somewhat depleting the sweep air of O_2. The air sweep flow can be limited to about 50% of the total combustion air and maintain near-maximum benefits of the sweep (i.e., increased CO_2 driving force). [12] Thus the project team has chosen to limit the sweep air to feeding only the secondary air flow, corresponding to roughly 76% of the combustion air.

2.5.3 MTR CO_2 Membrane Design Parameters

The module vessels that MTR currently uses for natural gas CO_2 removal are 26 ft long x 5 ft diameter cylinders. These high pressure vessels weigh 15 tons fully loaded with 2,600 m^2 of membrane. Adapting this technology for low-pressure flue gas, MTR estimates a weight of 7 tons including skid supports. These weights and dimensions have been used for preliminary layouts. Ultimately, MTR will look at redesigning module vessels for flue gas. One design being considered now is rectangular modules where a 1m x1m x1m box would contain 1,000 m^2 and weigh less than 2 tons. These rectangular modules could be easily stacked to increase the packing density and thereby reduce the footprint. [13] This information is summarized in Exhibit 2-9. The required area by module is presented in Exhibit 2-10.

Exhibit 2-9
Module Vessel Design Parameters

Module	Area (m^2)	Dimensions	Weight (fully loaded)	Notes
Current Module Vessels for **NG CO_2** removal	2,600 m^2	26'L x 5' D	Hi P App: 15 tons Lo P App: 8 tons (including skids)	For comparison only.
Historical Multi-Tube CO_2 Module System	5,600 m^2	8'Hx8'Wx15L'	NA	Ref: [14] Used in 2010 1st Quarterly Report
Prelimary Cylindrical Vessel Design: for Flue Gas	6,000 m^2	25'L x 5'D	Loaded vessel Weight: 7 tons	Ref. [15]. For use in analysis.
Possible Compact Design: Module Vessels for Flue Gas	1,000 m^2	1m x 1m x 1m	<2 tons	Easily stacked boxes. To be refined in future.

Reference: [13].
Note: Grey background indicates data is provided for information only. Non-greyed data is the design basis.

Exhibit 2-10
Required Membrane Area

Case	Module A Area (m^2)	Module B Area (m^2)	Module C Area (m^2)	Notes
MTR-1 (2 bar feed P for Module A)	217,000	325,000	4,275 m^2	Subject to change

Reference: [16]

2.6 TECHNICAL MATURITY

This study is based on technology that is presently technically feasible, but not necessarily available as commercially offered equipment. Bringing the required equipment to the commercial market could require some development by an OEM. DOE/NETL does not require these costs to be reflected in the cost estimate. This study is based on the position that Non-Recurring Engineering (NRE) costs are not included in the cost estimate of this study.

Where equipment required or assumed for this retrofit application is not commercially available, such equipment is identified as such. Equipment in or near this category include the following:

- CO_2 Membrane. MTR provided membrane performance based on what they were achieving at lab scale in June 2010 and projecting as their standard membrane performance achievable by the end of 2010. [16]
- Vacuum Pump. Although the vacuum pump application is beyond the limits of some suppliers due to the size, gas composition, and/or required efficiency, the modeled pump is based on a commercial available model from MAN Turbo.
- Flue Gas Compressors. Although the FG compressor application is beyond the limits of some suppliers due to the size, and/or gas composition, the modeled pump is based on a commercial available model from Dresser Rand.
- Flue Gas Expander. WorleyParsons did not receive confirmation of commercial availability for this high volumetric flow, moderate temperature, low pressure expander.

However, there is no reason such an expander couldn't be developed for a commercial market. For this present analysis, an efficiency of 87% was utilized.

- Low Temperature Expander. The low temperature expander is a custom design item, and is common to cryogenic processes such as an air separation unit. WorleyParsons obtained budgetary cost estimates for a single expander, and a cost savings assuming the non-recurring engineering (NRE) costs could be excluded. The NRE costs were excluded in the present analysis.

The CO_2 membrane for the pulverized coal plant is a novel application. No commercial scale units are in operation. A demonstration unit is in testing at the APS Cholla Power plant in Arizona. The cost of the membrane units will be priced to exclude the NRE costs.

2.7 MISCELLANEOUS

2.7.1 Water

The design water condition and characteristics will not have a significant impact on the retrofit application. The design basis for the cooling water, condensate, raw water and potable water are presented in Appendix D, Section 2.8.

2.7.2 Capacity Factor

Per Reference [1], the expected annual operating time is 7,446 hr/yr and is consistent with an 85% capacity factor.

2.7.3 Balance of Plant

The balance of plant design basis for water, waste water, and plant distribution voltage are presented in Appendix D, Section 2.14.

2.8 BASIS OF MEA CASE FOR COMPARISON

The base amine CO_2 scrubbing case against which the MTR CO_2 membrane option will be compared is the Reference [1] Case 1 Option that is based on 90% CO_2 capture with a solvent regeneration energy of 1550 Btu/lb CO_2. [17] This is the state-of-the-art (SOA) advanced amine case designed in 2006, which is 34% more efficient that the state-of-the-art amine case from 2000. [1] WorleyParsons has escalated the cost information in the reference document for Case 1. EPRI will utilize the original performance information and the escalated cost data for a performance and economic comparison to the MTR CO_2 membrane option.

It has been suggested that the Reference Study Case 1a (90% CO_2 capture with a regeneration energy of 1200 Btu/lb CO_2) could be used as a future State-of-the-Art amine cycle benchmark for the MTR CO_2 membrane case. The comparison to this Case 1a amine case is worthy of the following notes:

- Case 1a is a sensitivity case and is not one of the fully costed cases in the reference study. Since the performance is simply presumed and not based on known technological advances, the capital costs were not developed in the study. In the reference study, Case 1a is based on the presumed performance of 1200 Btu/lb CO_2 regeneration energy and an identical capital cost as Case 1. (Ref Study [1], p150, 5th bullet) This is clearly an aggressive comparison, since the performance improvements will cost more to implement.

- Possible process improvements that could contribute to the envisioned performance include an increased use of heat exchangers, increased solvent concentration, added inhibitors, advanced amines, mechanical vapor recompression (MVR) or other performance improving features. All improvements may increase auxiliary electric power requirement and would likely increase the capital cost.
- Since the cost for Case 1a is not developed in the reference study, and since the Case 1 cost presumed applicable to Case 1a is clearly low for Case 1a, care should be used in the comparison of Case 1a to the MTR CO_2 membrane.
- It has also been suggested that the advanced SOA amine case utilize a regeneration energy of something between 1200 to 1300 Btu//lb CO_2. The selection of the precise regeneration number is not relevant to WorleyParsons scope and is deferred to MTR and EPRI.

Since WorleyParsons is tasked with escalating the MEA cost in the reference study, and since the Case 1a cost is presumed to be equivalent to that of Case 1, WorleyParsons has simply escalated the Case 1 cost. EPRI will perform the appropriate sensitivity study and document it with the appropriate caveats.

3. CONESVILLE UNIT 5 DESCRIPTION AND MAJOR COMPONENTS SPECIFICATIONS

This section presents selected information of the Unit 5 existing major systems. The information characterizes possible limitations of the existing equipment and their suitability for the CO_2 membrane retrofit. That is, the CO_2 membrane retrofit design case presented in this study have been developed within the limitations of the existing equipment. The ability of the existing equipment to support operation of the retrofitted unit is evaluated to determine whether additional equipment is required.

3.1 OVERVIEW

Conesville Unit 5 is a subcritical pressure (2,400 psig/1,000°F/1,000°F) Rankine cycle pulverized coal fired plant. Unit 5 was commissioned in 1976, and it is one of six coal fired steam plants located on the Conesville site, which has a total installed generating capacity of ~2,080 MWe. The fuel utilized is bituminous coal from the state of Ohio. The flue gas leaving the steam generator system is cleaned of PM in an ESP and of SO_2 in a lime-based FGD system before being discharged to the atmosphere.

3.2 STEAM GENERATOR

Unit 5 utilizes a balanced draft, controlled circulation type, pulverized coal boiler. The Unit 5 boiler is equipped with a four-corner tilting/tangential concentric firing combustion system, comprised of five burner elevations. Pulverized coal for each elevation of burners is supplied by a single RP-903 pulverizer. A few years ago the Unit 5 boiler combustion system was retrofitted with a Separated Overfire Air (SOFA) low NOx system. The SOFA system reportedly does not significantly affect boiler design operation [6].

A summary of the existing Unit 5 air-fired steam generator technical information as specified in the original contract with Combustion Engineering is provided in Exhibit 3-1 [18].

Exhibit 3-1
Conesville Unit 5 Steam Generator Predicted Performance – Contract Fuel Basis

Parameter	Units	Control Load (50%)	Guaranteed (100%)	MCR (Note A) (VWO & 105%OP)
Manufacturer		Combustion Engineering, Inc.		
Commercial Operation	starting year	1976		
Superheater Flow	lb/hr	1,416,022	2,832,044	3,131,619
Superheater Pressure	psig	2,425	2,500	2,620
Superheater Temperature	°F	1,005	1,005	1,005
Reheater Flow	lb/hr	1,277,691	2,512,295	2,767,764
Reheater Inlet Temperature	°F	554	646	652
Reheater Inlet Pressure	psig	285	575	635
Reheater Outlet Temperature	°F	1,005	1,005	1,005
Reheater Outlet Pressure	psig	269	543	601
Feedwater Temperature	°F	416	482	493
Boiler Efficiency (Note B)	%	89.62	88.65	88.43

Notes: [18]

A. MCR – Maximum Continuous Rating
B. Based on contract fuel: Midwestern Bituminous Coal,
Moist=8.45%, Volatile= 36.11%, Fixed Carbon= 40.49%, Ash= 14.96%, HHV= 10,979 Btu/lb.

The predicted performance presented in Exhibit 3-1 is based on the contract fuel and conditions. For the present study, it is important to match the reference MEA CO_2 capture study [1] boiler fuel, performance, and operating conditions. To facilitate comparison to the reference MEA CO_2 capture study, the steam generator is assumed to operate at a VWO flow of 3,131,619 lb/h with 5% overpressure. The reference amine study presents a base case boiler efficiency of 88.13% based on the Midwestern coal presented in Exhibit 2-4. The carbon dioxide emission for the base case in the reference case is 866,156 lb/hr. The steam coil air heater is assumed to be off. The burner tilt is assumed to be at -10 degrees for the VWO 5% OP condition.

Additional technical information regarding the existing steam generator, including the elevation cross section, air, flue gas, feedwater and steam temperatures, pressures, pressure drops, and coal drying strategy can be found in the design basis document, Appendix D, Section 2.5.

3.3 AIR / FLUE GAS HANDLING SYSTEMS

Conesville Unit 5 air handling system is comprised of two Primary Air (PA) fans, two Forced Draft (FD) fans, and a Lungstrom-type tri-sector air preheater. A summary of the Unit 5 air/gas handling system design assumptions is presented in Exhibit 3-2.

Exhibit 3-2
Air / Flue Gas System Design Conditions

Parameter	Units	Value
Primary / Secondary Air Split (to FD Fans)	fractional	0.241 / 0.759
Furnace Pressure	in wg	-0.5
Excess Air Above Stoichoimetric	% wt	15
Infiltration air, based on total O_2 Requirement (infiltration air / 115% air)	%	5.3

The purpose of the PA fans is to supply combustion air to each of the five boiler pulverizers. Unit 5 is equipped with two PA fans each sized for 100% of total air flow demand. PA fan design information is summarized in Exhibit 3-3.

Exhibit 3-3
Primary Air Fan Design Data

	Units	Value
Fan Manufacturer		Buffalo Forge
Design Point Flow Rate	lb/hr	720,000
Design Point Pressure	in H_2O	46.15
Corresponding Fan Speed	rpm	1,180
Fan Efficiency at Design Point	%	75
Fan Energy Use at Design Point	hp	1,595
Fan Motor Manufacturer		Electric Machinery Mfg. Company
Fan Motor Capacity	hp	1,750
Fan Motor Voltage	V	4,000

Unit 5 is equipped with two FD fans (also referred to as a secondary air fan, or SA fan). The purpose of the FD fans is to supply the balance of air to the boiler windboxes required for complete combustion. FD fan design information is presented in Exhibit 3-4.

Exhibit 3-4
Forced Air Fan Design Data

Parameter	Units	Value
Fan Manufacturer		Westinghouse Electric Corporation
Design Point Flow rate	lb/hr	1,940,000
Design Point Pressure	in H_2O	26
Corresponding Fan Speed	rpm	880
Fan Efficiency at Design Point	%	88.2
Fan Energy Use at Design Point	hp	2,290
Fan Motor Manufacturer		Electric Machinery Mfg. Company
Fan Motor Capacity	hp	2,500
Fan Motor Voltage	V	4,000

The air preheater heats primary and secondary air streams while reducing flue gas temperature prior to entering into the ESP.

Exhibit 3-5
Air Preheater Design Data

Item	Units	Value
Manufacturer		The Air Preheater Company, Inc.
Type		Ljungström, rotary regenerative, vertical shaft, counter flow, tri-sector,
Convective Heating Surface	ft^2	440,600

Major components of the flue gas system are the ESP and the induced draft (ID) fans. The ID fans are used to maintain constant pressure in the furnace. Unit 5 is equipped with two ID fans, which are presumed to be designed as 2x50% of flue gas system total design flow rate, but typically operate as 1x100%. ID fan design information is presented in Exhibit 3-6.

Exhibit 3-6
Induced Draft Fan Design Data

Parameter	Units	Value
Fan Manufacturer		Green Fuel Economizer Company
Design Point Flow Rate	lb/hr	2,765,000
Design Point Pressure	in H_2O	46
Corresponding Fan Speed	rpm	880
Fan Efficiency at Design Point	%	87.5
Fan Energy Use at Design Point	hp	7,786
Fan Motor Manufacturer		Electric Machinery Mfg. Company
Fan Motor Capacity	hp	8,000
Fan Motor Voltage	V	4,000

The Unit 5 ESP is a rigid-frame precipitator comprised of four gas passages each equipped with eight collecting-electrode assemblies. The Unit 5 ESP design information summary is presented in Exhibit 3-7. The dust collection efficiency utilized in this analysis is 99.65%.

Exhibit 3-7
Electrostatic Precipitator Performance Data

Parameter	Units	Value
Guaranteed Minimum Collection Efficiency (Note A)	%	99.65
Inlet Dust Loading, 225-300°F	grains/acf	2.05 – 6.8
Outlet Dust Loading	grains/acf	0.01
Maximum Gas Velocity Through the Unit (Note A)	fps	5.5
Design Flue Gas Mass Flow Rate	lb/hr	4,440,000
Design Flue Gas Volume Flow Rate	acfm	1,406,844

Notes:
A. At design conditions

3.4 FLUE GAS DESULFURIZATION SYSTEM

The existing Conesville Unit 5 is equipped with a wet lime based flue gas desulfurization (FGD) system to control SO_2 emissions as shown in Exhibit 3-8. The FGD system is of the natural oxidation type and generates primarily calcium sulfate/sulfite waste solids for disposal. The FGD system is located downstream from an ID fan. A summary of the existing FGD system performance is presented in Exhibit 3-9.

Exhibit 3-8
Existing FGD System Process Flow Diagram

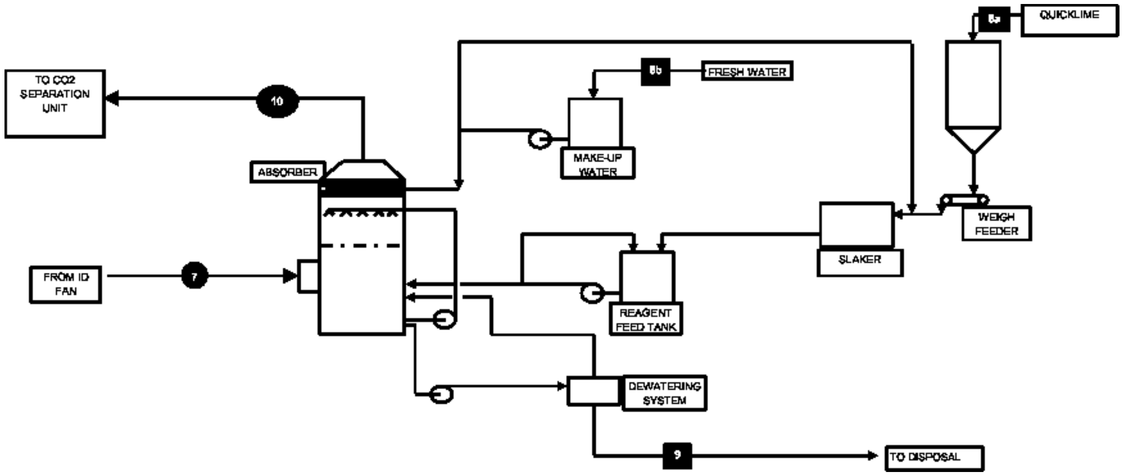


Exhibit 3-9
Existing FGD System Performance Data

Parameter	Units	Value	
Ca/S	Mole Ratio	1.04	
Solids	wt%	20	
Liquid/Gas (L/G) Ratio	gpm/1000 acfm	55	
Oxidation Oxygen/SO_2 Removal Ratio	mole ratio	2.3	
Oxidation Air Pressure	psig	0.45	
FGD Outlet Pressure	in wg	15.2	
FGD Outlet Temperature	°F	135.0	
CO_2/SO_2 Mole Ratio		63	
SO_2 Removal Efficiency (Absorber)	%	97.4%	
SO_2 Removal Efficiency (System)	%	94.9%[b]	
Flue Gas Composition		***Absorber Inlet***	***Absorber Outlet***
O_2	mol/hr	4,469	4,461
N_2	mol/hr	105,018	105,018
H_2O	mol/hr	12,853	24,228
CO_2	mol/hr	19,743	19,720
SO_2	mol/hr	315	16
O_2	vol%	3.14%	2.91%
N_2	vol%	73.74%	68.44%
H_2O	vol%	9.03%	15.79%
CO_2	vol%	13.86%	12.85%
SO_2	ppmv	2,212	104

A major design criterion determining the required FGD system SO_2 removal efficiency is the site environmental requirements. Upon conversion to the CO_2 membrane capture, controlling SO_2 with the existing FGD system for environmental purposes may not be required as most of the sulfur compounds (SO_2 and SO_3) would be co-sequestered with the CO_2.[c]

Over the course of the project, the project team considered whether FGD operation should continue with just water for PM mitigation and flue gas cooling, or with the normal lime slurry to reduce the SO_2 levels going into the CO_2 system. Operation of the existing FGD with lime was

[b] The absorber removal efficiency of 94.9% is known to now be better since the 2.5% bypass has subsequently been removed. Nevertheless, this analysis will be performed on a consistent basis with the Reference [1] study, which utilized the 2.5% bypass.

[c] SO_2 is twice as permeable as CO_2, so at 90% CO_2 capture, more than 90% of SO_2 will permeate and be captured by the membrane system. In addition, SO_3 would also permeate through the CO_2 membrane.

judged to be prudent simply to avoid having to deal with the SO_3 that would come out of the CO_2 stream during compression inter-cooling and water condensation. The lime will neutralize the H_2SO_4 that would form in either the FGD or the CO_2 compression system. It is therefore decided to continue normal operation of the FGD system with the lime slurry. This approach will also develop a sweet CO_2 product which may be required depending on the oil field being considered for EOR. In addition to sulfur reduction, continued operation of the FGD system will precondition the flue gas going to the membrane system by providing additional particulate matter removal and flue gas cooling.

3.5 STEAM TURBINE GENERATOR

The Unit 5 steam-turbine generator is a tandem compound machine, with high pressure (HP), intermediate pressure (IP) and low pressure (LP) sections that drive a single 3,600 rpm hydrogen cooled generator. The LP turbine is a double flow machine exhausting downward into the condenser. A summary of the existing Conesville Unit 5 steam turbine nameplate information is provided in Exhibit 3-10 and generator information in Exhibit 3-11 [19].

Exhibit 3-10
Conesville Unit 5 Steam Turbine Nameplate Data

Parameter	Units	Value
Manufacturer		Westinghouse
Maximum Output (implied PF=0.91)	kWe	448,759
Throttle Steam Pressure	psig	2,400
Throttle Steam Temperature	°F	1,005
Reheat Temperature	°F	1,005
Exhaust Pressure	in HgA	2.5

Exhibit 3-11
Conesville Unit 5 Generator Nameplate Data

Parameter	Units	Value		
Manufacturer		Westinghouse		
Voltage	volt	24,000		
Hydrogen Pressure	psig	30	45	60
Max. Apparent Power	kVA	395,000	445,000	493,280
Power Factor		0.90	0.90	0.90
Max. Real Power	kWe	355,500	396,000	443,952

As discussed in the Reference MEA study [1], the base case steam turbine cycle for this analysis is based on a valves wide open (VWO), 5% overpressure (OP) case utilizing a condenser backpressure of 6.35 cm Hga (2.5 in Hga), no steam extraction for steam coil air heaters, and a reheat de-superheating spray of approximately 85,850 lb/hr. As documented in the reference study, the base case was adjusted so the steam turbine will be based on a swallowing capacity of 3,131,619 lb/h at 5% OP, and the generator efficiency was adjusted slightly to facilitate comparison to past studies by exactly matching the gross generation of 463,478 kWe. The base case steam turbine cycle heat and mass balance as utilized in this analysis is presented in Exhibit 3-14.

No modifications are expected for the existing steam turbine generator, although an additional extraction from the IP – LP cross over piping is expected to provide steam for regeneration of the tri-ethylene glycol system required for the CO_2 drying process.

Key parameters for the Conesville Unit 5 steam turbine generator as analyzed for consistency to the reference MEA study are presented in Exhibit 3-12. Additional data can be found in heat and mass balance provided in Exhibit 3-14.

Exhibit 3-12
Conesville Unit 5 Steam Turbine Generator – As Analyzed

Parameter	Units	Value
Main Steam Flow	Lb/hr	3,131,619
Throttle Steam Pressure (105% overpressure)	psig	2,520
Throttle Steam Temperature	°F	1,000
Hot Reheat (HRH) Steam Flow, outlet	Lb/hr	2,851,907
HRH Steam Pressure (105% overpressure)	psig	608
HRH Steam Temperature	°F	1000
Reheat De-superheating Flow	Lb/hr	85,877
Implied Power Factor	%	93.96%[d]
Gross Generation	kW	463,478

3.6 UNIT 5 POWER OUTPUT RATING

Conesville Unit 5 uses a single reheat 2,400 psig/1,000°F/1,000°F Rankine cycle. Power rating of a Rankine cycle unit can be defined by several operating modes including:

- 100% throttle flow (typically a guarantee condition). At this design point, the steam turbine generator typically operates at the condition that the turbine control valve is not wide open. This permits the steam turbine generator control system to modulate; and
- Combined 5% Overpressure and Valves Wide Open mode (5%OP/VWO). This combined condition is typically expected to be used on relatively infrequent occasions to support severe sustained demand conditions.

[d] The implied power factor of 93.96% is based on the gross generation of 463,478 kW, and the generator MVA of 493,280 kVA for 60 psig H_2 pressure.

A summary of Unit 5 major operating parameters at the 100% throttle flow mode [20] and the 5%OP/VWO condition [21] is presented in Exhibit 3-13.

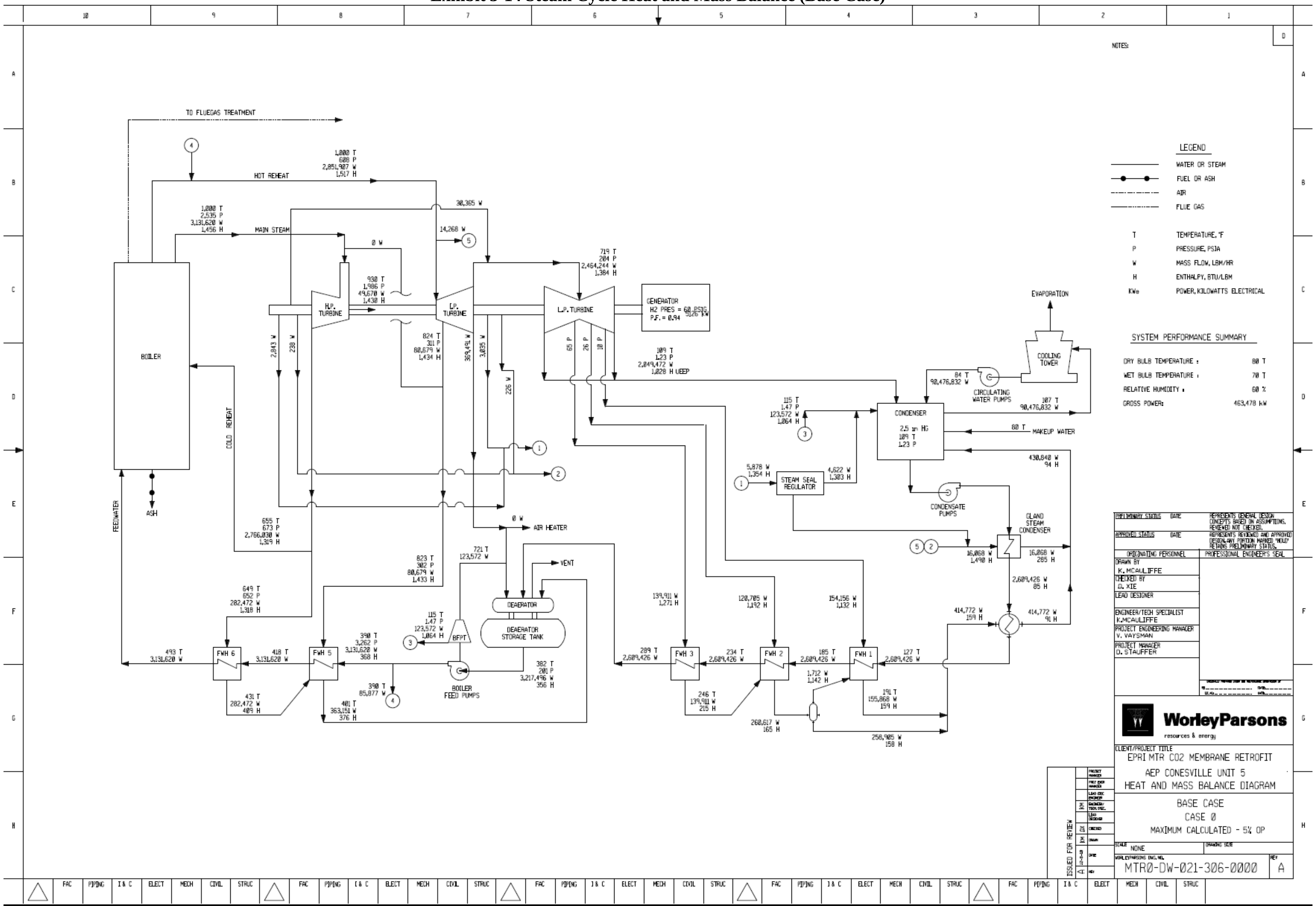
Exhibit 3-13
Unit 5 Gross Power Output Rating

Parameter	Units	100% Throttle [22]	5% OP/VWO [1]
Throttle Steam Flow	lb/hr	2,832,044	3,131,619
Throttle Steam Pressure	psig	2,400	2,520
Throttle Steam Temperature	°F	1,000	1,000
Condenser Pressure	in HgA	2.5	2.5
Generator Gross output	kWe	412,852	463,478
Power Factor		0.90	0.9398
Hydrogen Pressure	psig	60	60

At the combined 5%OP/VWO condition, Unit 5 gross output exceeds its output at the 100% throttle flow condition by almost 51 MWe, or 12.3%. However, Rankine cycle based plants typically are not designed to operate continuously at 5% OP/WVO condition for an extended period of time. This operating point is selected as the basis for this retrofit analysis in order to be consistent with the reference study performed for the MEA study.

The existing Unit 5 steam cycle heat and mass balance for the VWO, 5% OP condition is presented in Exhibit 3-14.

Exhibit 3-14 Steam Cycle Heat and Mass Balance (Base Case)



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3.7 STEAM CYCLE HEAT SINK

The existing Unit 5 utilizes mechanical draft evaporative cooling towers and a circulating water system as a heat sink for the steam cycle and auxiliary heat loads. The Conesville Unit 5 is equipped with two 5-cell cooling towers (Plant designation 5A and 5B). The cooling towers were manufactured by Balcke-Durr and installed in 1999. In addition to the main condenser load, heat from the auxiliary cooling water system and the bearing water system is rejected through the cooling towers. A summary of the existing cooling tower design technical information is presented in Exhibit 3-15 [23; 24].

Exhibit 3-15
Existing Cooling System Design Data

Parameter	Units	Value
Wet Bulb Temperature	°F	75.0
Approach Temperature	°F	16.2
Cold Water Temperature	°F	91.2
Range	°F	25.0
Hot Water Temperature	°F	116.2
Circulating Water Flow (two 5-cell towers)	gpm	174,000
Fan Brake Horsepower (ea)	hp	200

Unit 5 circulating water system is equipped with two 50% capacity circulating water pumps that circulate cooling water between the cooling tower basin and main condenser. Each pump is a single stage vertical wet pit centrifugal unit with main characteristics as presented in Exhibit 3-16.

Exhibit 3-16
Circulating Water Pumps Design Data

Parameter	Units	Value
Pump Manufacturer		Ingersoll-Rand Corporation
Design Point Flow Rate	gpm	87,000
Total Discharge Head	ft H_2O	95
Corresponding Pump Speed	rpm	440
Pump Efficiency at Design Point	%	89
Pump Energy Use at Design Point	hp	2,345
Pump Motor Manufacturer		Electric Machinery Mfg. Company
Pump Motor Capacity	hp	2,500
Pump Motor Voltage	V	4,000

The cooling tower makeup water characteristics assumed in the study are presented in Exhibit 3-17. This makeup water quality will allow cooling tower operation with 4 cycles of concentration of dissolved solids in the circulating water.

Exhibit 3-17
Makeup Water Characteristics

Constituent	Formula	Units	Design Value
Calcium	Ca	mg/l	75
Magnesium	Mg	mg/l	16
Potassium	K	mg/l	3
Sodium	Na	mg/l	20
Bicarbonates	HCO_3	mg/l	240
Chlorides	Cl	mg/l	25
Silica	SiO_2	mg/l	4
Sulfates	SO_4	mg/l	58
Nitrate	NO_3	mg/l	7
Total Dissolved Solids		mg/l	460
Total Organic Carbon		mg/l	3
Temperature		°F	80
pH			8.0

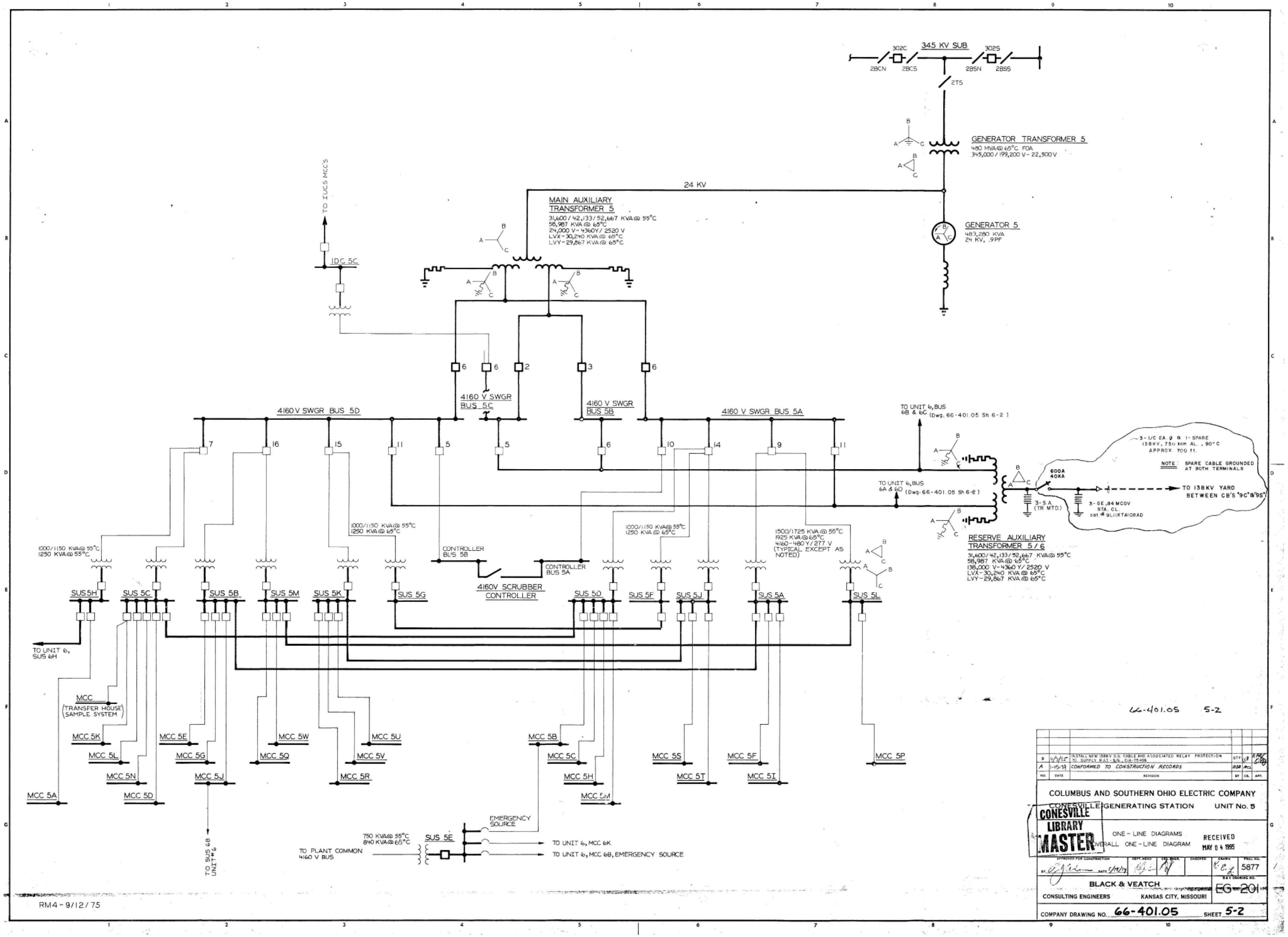
3.8 ELECTRICAL SYSTEM

Assumed plant voltage distribution is provided in Exhibit 3-18, and a one-line diagram of the existing Unit 5 electrical system is presented in Exhibit 3-19.

Exhibit 3-18 Assumed Plant Voltage Distribution

Load	Voltage
Motors Below 1 hp	110/220
Motors 250 hp and Below	480
Motors Above 250 hp	4,160
Motors Above 5,000 hp	13,800
Steam Turbine Generators	24,000

Exhibit 3-19 Unit 5 Electrical System One-Line Diagram



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4. CO_2 MEMBRANE RETROFIT

This section provides a description of the retrofit, new systems, modifications to existing systems, the plant layout, performance, and system integration.

Supplementing this section are the heat and mass balances in Appendix A, the Block Flow Diagram (BFD) and Process Flow Diagrams (PFD) in Appendix B, and the major equipment list in Appendix C.

4.1 BFD & RETROFIT DESCRIPTION

The retrofit of Conesville Unit 5 with the MTR CO_2 membrane system will require the addition of new systems and modifications of the operation of the existing systems.

The following major systems will be added to Unit 5 as a part of the CO_2 membrane retrofit:

- Modules A, B, and C of the CO_2 membrane enclosures complete with a gas distribution system and structural supports
- Flue Gas Direct Contact Cooler system
- Flue Gas compressor and Vacuum pumps
- Secondary Air Booster fan
- Flue Gas Expander
- CO_2 Purification and compression system comprised of CO_2 compression/dehydration, chiller and distillation systems.
- Flue Heat exchangers with glycol circulating systems.
- Process cooling system comprised of new auxiliary cooling tower, circulating water system and water treatment to accommodate additional process cooling loads

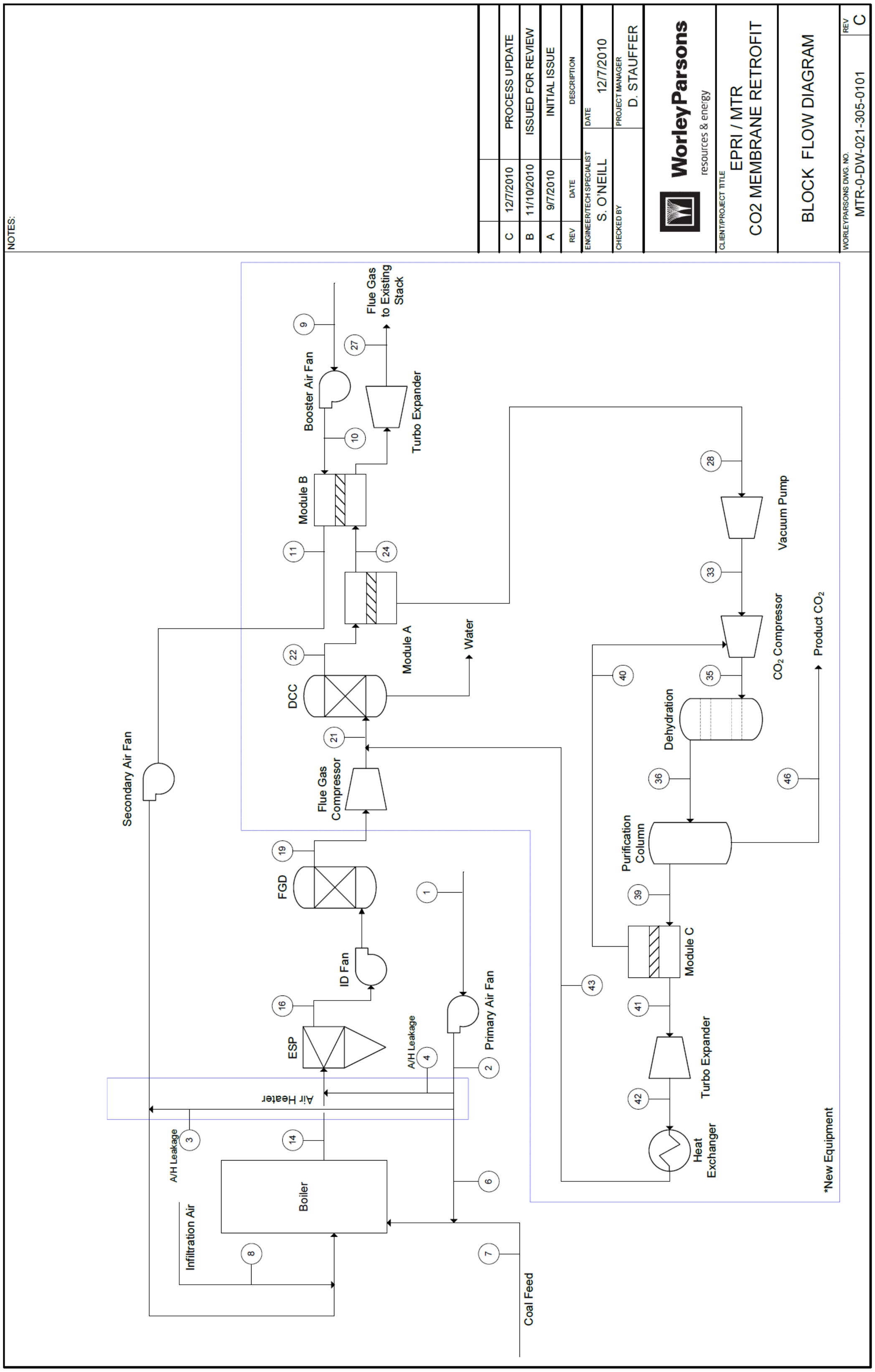
Modifications to the existing Unit 5 systems are expected to include:

- Modifications to the secondary air ducts
- Modifications to the flue gas ducts
- Changes in operation of the secondary air and flue gas handling systems

The new systems and modifications to the operation of the existing Unit 5 systems are described in the following sections. The retrofitted unit block flow diagram (BFD) is presented in Exhibit 4-1.

The Process Flow Diagrams (PFDs) are presented in Appendix B. Corresponding heat and mass balance tables are presented in Appendix A.

Exhibit 4-1 Block Flow Diagram of the Retrofitted Unit 5



4.2 NEW SYSTEMS DESCRIPTION

This section provides a description of the new systems added as part of the CO_2 membrane retrofit. A major equipment list which characterizes this added equipment is presented in Appendix C.

Flue gas at the Conesville plant will encounter the first piece of newly installed equipment in the ductwork between the ID fans and the FGD absorber. Finned heat exchanger tubes are to be installed in the duct work to recover energy from the hot flue gas into a circulating stream of glycol. Energy collected from the flue gas after the ID fan is carried by the glycol to the exit of the counterflow MTR module, where a second set of finned tubes heat the pressurized flue gas before the flue gas expander. This transfer of energy allows the expander to achieve a greater power output and will maintain additional thermal buoyancy in the flue gas exiting the stack. The lower flue gas temperature entering the FGD absorber also helps to reduce the amount of water which is evaporated.

Flue gas stripped of SO_2 in the FGD absorber is fed to four compressors operating in parallel. These flue gas compressors raise the pressure up to the 2 bar design pressure of the MTR membranes. The pressurized flue gas is combined with retentate from the MTR crossflow module C and cooled in a direct contact cooler vessel. In the vessel, which is similar to an FGD absorber, cold water is sprayed over the gas to lower the temperature. Cold, pressurized flue gas is distributed by a header system to the banks of MTR crossflow module A membranes.

MTR's membranes capture CO_2 by using partial pressure as a driving force across a selective barrier material. The selectivity of the material allows a greater percentage of the CO_2 to preferentially permeate the membrane while those compounds which would be impurities in a CO_2 product preferentially pass through as a retentate stream. A vacuum on the permeate side of these crossflow membranes provides additional pressure gradient to drive the CO_2 capture.

Retentate from the crossflow module A membranes is distributed through banks of MTR counterflow module B membranes. In these membranes the CO_2 permeates from the flue gas into the boiler's secondary air. This membrane creates a CO_2 recirculation loop within the plant to ensure that the desired 90% CO_2 capture level is achieved. A booster air fan will be installed to drive the secondary air through the banks of module B membranes. The CO_2 depleted flue gas which exits the module B membranes is at a pressure greater than what is required to ensure proper dispersion through the plant stack.

The second set of finned tube heat exchanger tubes are installed in the ductwork between the crossflow module B membranes and a flue gas expander. These tubes transfer energy from the recirculating glycol into the flue gas, raising the temperature of the gas. A single stage expander recovers energy from the hot pressurized gas as the pressure is reduced from the operating pressure of the membrane modules to the pressure required to dispel the gas through the stack.

Dry vacuum pumps are utilized to maintain vacuum on the permeate side of the crossflow module A membranes. Liquid ring vacuum pumps which are used at power plants to maintain the vacuum in the condenser are not well suited to this membrane service. The CO_2 and SO_2 are water soluble gases and water is used in high volumes as a sealant in a liquid ring vacuum pump. Dissolution of CO_2 into the water reduces the systems capture percentage and is undesirable in this application. Auxiliary power consumption by liquid ring pumps is prohibitive to the process in addition to the CO_2 losses through the water. Alternately to liquid ring vacuum pumps are dry type compressors which are used to achieve large volumes of vacuum in the pulp and paper industry and are better suited to MTR's process.

Dry compressors operate more efficiently and consume less power. The vacuum pumps for this large volume service would be designed with a combination of axial and radial flow stages to achieve the desired level of vacuum. Two large vacuum pumps are required by the process to capture permeate from MTR crossflow module A. The CO_2 rich permeate from MTR's counterflow module A also contains water and SO_2 which have an influence on the vacuum pump materials of construction. At the discharge of the vacuum pump system the CO_2 rich gas is cooled, water is removed, and the gas is piped to machines which will perform further compression.

Three multistage compressors operating in parallel will be required to process the volume of gas present following the vacuum pumps. Permeate from MTR's crossflow module C is introduced into one of the later stages of compression. Circulating water from a new cooling tower will be supplied to coolers after the compression stages to remove heat which is generated by the compression process. Efficient intercooling reduces the auxiliary power consumption of the machines. Pressurized CO_2 rich gas exiting the compressors must be dried and purified before being pumped to the final boundary limit pressure.

A Tri-Ethylene Glycol (TEG) drying system is installed following the compression system to remove moisture which was not knocked out in the compression process. The TEG is a temperature swing chemical absorption system in which lean and rich TEG solvent is circulated between the regenerator and absorber. For this application where less than 50 ppmv of H_2O in the product gas is required, a high dew point depression TEG dehydration process has been utilized.

Dry CO_2 rich gas still contains impurities which exceed the values specified in Exhibit 2-7 and must be further treated.

Purification of the CO_2 rich gas to produce CO_2 which meets the specified requirements is done utilizing a low-temperature partial condensation process integrated with a distillation column. Cooling water has already been utilized to reduce the temperature and other heat sinks must be used. The gas exiting the CO_2 drying system is cooled down to the necessary temperature in two stages. The relatively hot CO_2 rich gas leaving the drying system is used to meet the energy demands of the CO_2 stripping column reboiler in the first stage of the cooling at the same time eliminating a process steam demand. A chiller system based on evaporation of liquid propane is used to reduce the temperature further and partially condense the CO_2 rich gas.

Condensation of the CO_2 dominates at the design temperature and pressure of the gas condenser (HX3-H). Oxygen and nitrogen condense with the CO_2 in values that exceed the product specification. Impurities in the CO_2 are removed by processing the liquid mixture in a stripping column. As the impure liquid CO_2 cascades down the column, vapor which is generated in the reboiler travels upwards. The impurities preferentially fractionate into the vapor phase as it moves up and out of the column. A pure liquid CO_2 product which meets all specifications is drawn off of the bottom and pumped up to the final discharge pressure.

Overheads from the CO_2 stripping column contain a residual fraction of CO_2. MTR's counterflow module C recovers a portion of the CO_2 and returns it to the compression system. The pressure differential between the column overheads and the suction pressure of the compressor stage is used as the driving force for permeation. The retentate gas which passes through this counterflow module is at high pressure and still contains CO_2. Power is recovered from the membrane retentate through a low temperature expander. Expanded retentate is reintroduced into the flue gas prior to the direct contact cooler to give the overall system another chance to capture the recycled CO_2.

A chilling system which utilizes propane as the refrigerant is supplied to achieve the temperatures necessary to condense the CO_2 mixture. Gaseous propane is compressed up to a pressure which will facilitate condensation at a temperature which can be achieved by cooling water. The efficiency of the chilling system is increased by using the liquid propane to reject heat to the purification process. Expanded module C retentate, product CO_2, and stripping column overheads are all at temperatures lower than that of the liquid propane. Through heat exchange with those three gases the liquid propane can be sub-cooled. This process reduces the losses associated with reducing the pressure of the liquid propane. At the reduced pressure the liquid-vapor mixture of propane is sent to the CO_2 rich gas condenser where the liquid propane evaporates inducing condensation in the CO_2 rich gas.

A new cooling tower and auxiliary cooling water system will be installed to meet the new process cooling demands. A set of circulating water pumps will be installed for the new tower. The circulating water will service the vacuum pumps, multistage compressors, direct contact cooler, and propane compressor systems. Makeup water demand for the new cooling tower will be offset by collecting the condensate from the DCC and the compression process and pumping it to the cooling tower basin.

4.3 MODIFICATION TO EXISTING SYSTEMS

In retrofitting an MTR CO_2 capture system onto an existing power plant the existing systems may be required to operate at different conditions. Some pieces of equipment may be capable of performing their required duty at the new operating point while others may require modifications. This section details what effect an MTR CO_2 capture system retrofit has on the Conesville Unit 5 systems.

4.3.1 Steam Generator

The steam generator secondary air system is used as a sweep gas in the MTR counterflow module B to enhance CO_2 capture. This results in the vitiated air being fed into the secondary air system and an increased mass and volumetric flow rate through the system, as presented in Exhibit 4-2.

Exhibit 4-2
Changes in Secondary Air System Process

Parameter	Units	Base case	Retrofit Case
Secondary Air Flow Rate			
Mass	lb/hr	2,843,126	3,539,302
Volumetric	MMscfd	901	1,071
CO_2 Content	Mol%	0.03	9.10
O_2 Content	Mol%	20.52	17.36
Average Molecular Weight	g/mol	28.74	30.11

Alstom is the original equipment manufacturer (OEM) of the Conesville Unit 5 boiler. Alstom declined to provide an analysis of the MTR membrane system impact to the Conesville Unit 5 boiler. Presented in this is a section qualitative assessment of the impact of the CO_2 membrane retrofit on the existing steam generator system that is based on the following:

- WorleyParsons engineering judgment and knowledge of boiler performance and design
- The teleconference with Alstom earlier in the project [25]
- Utilization of a Boiler Performance Model in a generic way [26].

4.3.1.1 Secondary Air System

The increased mass flow through the secondary air systems will have the effect of increasing the pressure drop through the ductwork. The FD fan must now operate with a higher flow and produce a greater pressure differential. Conesville Unit 5 has two FD fans already built into their system. One fan alone is not capable of meeting the demands of the new operating point and as such both fans will need to run in parallel. There is a possibility that this operating scenario may be unstable due to the flatness of the fan system pressure curve. Oscillation of the discharge flow and pressure may occur. Installation of variable speed drives on the FD fans may stabilize their performance. Further investigation is required into the operating specifics of the FD fans. This report assumes that the fans would operate properly without the additional cost of retrofit motors. With both fans in operation, the secondary air is propelled through the air heater and into the boiler furnace.

4.3.1.2 Flue Gas Handling System

The ID fan system pulls the flue gas through the electrostatic precipitator once the flue gas has been cooled by the air heater. This system generates sufficient pressure to expel the flue gas through a flue gas desulphurization absorber and up the plant's stack. In the new arrangement this system must transport the gas to the inlet of a flue gas compression system. The mass flow that must be blown by the fans as well as the amount of pressure head they must generate is increased by the MTR retrofit. Two ID fans are installed similar to the FD fan system at Conesville. Due to the new operating flow and pressure, the operation of both fans will be required to operate to meet the systems demands.

4.3.1.3 Boiler Performance

Introduction into the boiler furnace of extra CO_2 and N_2 with the MTR secondary air sweep gas system is expected to reduce furnace temperature, and steam generation, while flue gas velocity is expected to increase. Heat absorption in the boiler will be shifted to the boiler back pass and superheated steam temperature would increase. An assumed solution is to tilt the burners down, shifting heat absorption to the furnace, and correcting both steam generation and steam temperature. Compensating for the temperature shift to the boiler's back pass surfaces (superheating, re-heating and economizing) requires sufficient temperature control margin in the existing tilts and de-superheating spray. The 20 degrees of available tilt (according to the previous study [1]) appears to be sufficient for temperature control. If not, some superheater, reheater, and economizer surfaces may be replaced with additional evaporative surfaces. Until a boiler vendor can perform a more detailed assessment, it assumed that the Conesville Unit 5 boiler is capable of accommodating the required steam generation and temperature control without modification. Thus, boiler duty is expected to be unaffected for a given steam mass flow, pressure and temperature, assuming that sufficient temperature control (tilts, & spray) is available.

The reduced O_2 content in the secondary air is likely to increase the unburned carbon (UBC) in the ash. The UBC affects the boiler efficiency and CO levels. The UBC may be mitigated through improving the fineness of the pulverized coal. This is analogous to the pulverizer modifications required for low NOx burner retrofits. A few years ago the Unit 5 boiler combustion system was retrofitted with a Separated Overfire Air (SOFA) low NOx system. It is assumed that this Low NOx retrofit included the pulverizer upgrade, and no capital cost penalties associated with the pulverizer upgrade will be included in the MTR membrane retrofit case.

It is estimated that the retrofit with the CO_2 membrane system is expected to result in a slightly greater boiler dry gas loss. Hence, efficiency of the retrofitted boiler is expected to be reduced by approximately 0.8%.

4.3.1.4 Combustion System

Since the O_2 content in the primary air and the primary air temperature will remain unchanged with the CO_2 membrane retrofit, no burner modifications are expected other than the re-tuning of the existing burners consistent with the tuning process originally performed for a Low NOx combustion system. For example, the secondary air distribution between boiler windboxes, closed coupled over-fire air (CCOFA) and SOFA may need to be adjusted with increased secondary air flow rate and reduced O_2 content. Assessment of the impact on windbox/CCOFA/SOFA operation requires detailed analysis and testing by the boiler vendor. SOFA operation may still be required to control NOx.

4.3.1.5 Boiler Back Pass

With an increased boiler flue gas flow rate there is a significant risk of increased erosion of the boiler back pass surfaces. Erosion increases exponentially with increased velocity and erosion also increases due to the higher molecular weight of the flue gas. Although no duct modifications are assumed in this study, a detailed analysis by a boiler OEM may recommend duct modifications to minimize erosion.

4.3.2 Effects on ESP

The PM collection efficiency of an ESP is dependent upon many factors, including volumetric flow, gas composition, temperature, contact time, resistivity, inter alia. Although parameters like resistivity and temperature are not expected to change significantly, the volumetric gas flow is expected to change by nearly 25%. This increased volumetric flow will decrease the contact time between the flue gas and ESP and will reduce the collection efficiency of the ESP. The extent of the collection efficiency change would be difficult to quantify and is outside of the scope of this evaluation.

Any increase in PM leaving the ESP would be more than compensated for by the downstream equipment. For example, the downstream FGD and DCC are expected to reduce the PM levels by approximately 75% and 40 to 50% respectively. Downstream of the DCC are the membrane modules. From a plant emission point of view, the emitted PM will be lower than pre-retrofit because of the newly added DCC and membrane system. The only PM concern is related to how much PM can pass to the membrane before causing premature fouling. The membrane demonstration project currently underway at the APS Cholla power plant should help to address the how much of an issue PM presents to the MTR membranes.

In summary, any anticipated performance change for the ESP is not expected to be an issue for the retrofit application, and no performance change has been accounted for.

4.3.3 Effects on FGD

The increased gas velocity going through the existing FGD will affect the SO_2 removal efficiency of the FGD. When below the flooding velocity, an increased velocity will tend to decrease the liquid droplet diameter, thereby increasing the contact area and increasing the FGD performance. However, when the gas velocity is above the flooding velocity, the liquid vapor interface will not be optimal, and the removal efficiency will decrease. [27]

The detailed evaluation of the FGD performance was outside of the scope of this study. However, from the emission compliance point of view, a decreased performance associated with the FGD will not be an issue since the membranes are permeable to both SO_2 and SO_3 and will therefore further reduce the sulfur levels in the flue gas by over 90% as discussed in Section 3.4.

In summary, any anticipated performance change for the FGD is not expected to be an issue for the retrofit application, and no performance change has been accounted for.

4.3.4 Effects on Steam Cycle

Electrical energy is the prime driver for the equipment which would be installed to capture CO_2 with MTR's membranes. As such, only a minor modification to the steam cycle is required as a result of the new systems. Low pressure steam is used to provide the heat required to dry the CO_2 before it is liquefied. A pipe would divert LP steam from the LP crossover to the drying unit. Steam condensate from the dryer is returned to the plant's condenser.

4.3.5 Electric System

In accordance with the plant voltage distribution (Exhibit 3-18), the loads of the auxiliary power distribution system are divided by their rating, i.e. power to loads larger than 250 HP is distributed at 4.16 kV, and power to loads of 5,000 HP and larger is distributed at 13.8 kV.

Power to loads rated at 250 HP and lower will be distributed at 480 V. The sizes of the new electric motors driving the flue gas compressors, vacuum pumps, and CO_2 compressors, and the chiller system exceed 5000 HP. The existing Unit 5 auxiliary power distribution system is comprised of 4.16 kV medium and 480 V low voltage systems. (Exhibit 3-19) Thus, the scope of modifications to the plant electrical system is expected to include addition of a new medium voltage 13.8 kV system to support operation of motors larger than 5000 HP. The new flue gas expander generator will be also connected to the new 13.8 kV system.

In addition, the existing 4.16 kV and 480 V systems will be modified to support operation of a CO_2 pump, a new process cooling system, a CO_2 dehydration system, and the recycle gas expander generator.

The scope of modifications of the auxiliary power distribution system is envisioned to include the addition of segregated-phase bus ducts, switchgear and control equipment, service transformers, generator equipment, station service equipment, conduit and cable trays, wire, and cable with all required foundations, and standby equipment.

Larger motors (such as the flue gas, vacuum and CO_2 compressors) will be equipped with a reduced voltage start up system[5].

4.4 NEW SYSTEMS / EQUIPMENT LAYOUT

The availability of space is a factor in the installation of MTR's CO_2 capture system. Limited space is available in the vicinity of the unit being retrofitted. Since the system would be negatively affected by gas pressure drops if the equipment was far removed from the existing boiler structures, a multistory building is designed to house the membranes adjacent to the Unit 5 FGD absorber vessels.

Large rotating equipment is most effectively installed at ground level. Ground level installation allows for the construction of rigid foundations capable of supporting the weight and vibrations of the rotating equipment. Membranes are installed on a floor above the various compressors in order to minimize ductwork and pressure drop.

Captured dilute CO_2 gas is ducted to a separate area of the plant for processing. Sufficient open land is available to facilitate the outdoor installation of the CO_2 compression and purification systems. A new cooling tower capable of servicing all of the new equipment is co-located with the CO_2 compression and purification equipment.

The proposed layout for the retrofit application is shown is Exhibit 4-3.

[5] Reduced voltage starting allows a motor to start in an unloaded mode, and reduces the current requirement by eliminating "inrush" (the apparent short circuit across the non-spinning windings). Inrush current level can exceed the normal current by a factor of 6 to 8.

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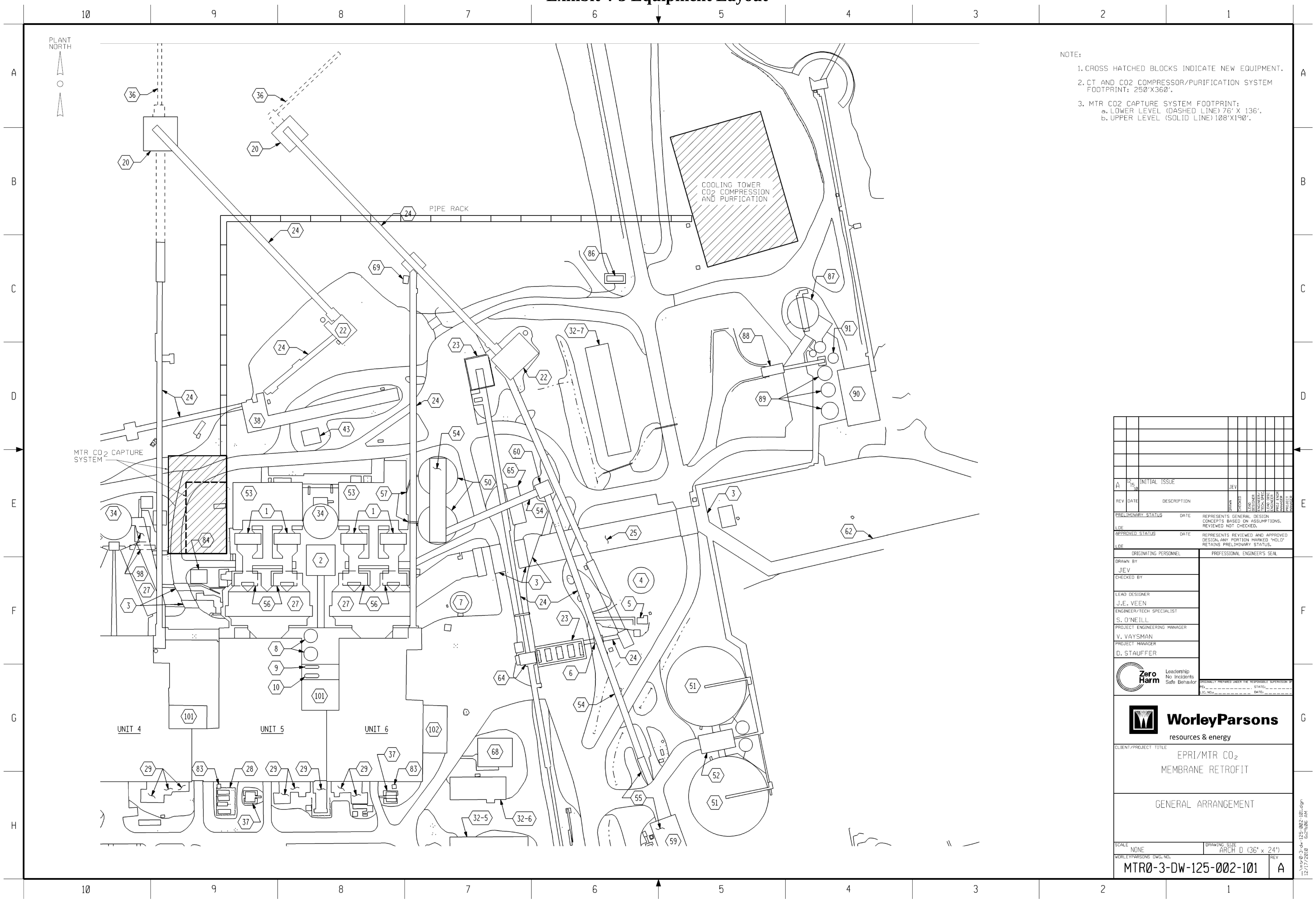
Exhibit 4-3 Equipment Layout



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4.5 RETROFIT PERFORMANCE

4.5.1 Gross, Net Power Generation and itemized Auxiliary Load

With the installation of new equipment and the change in the operating conditions of the existing plant, the net power generated by the plant is reduced. The retrofitted Unit 5 performance is summarized in Exhibit 4-4.

Exhibit 4-4
Retrofitted Unit 5 Performance Summary

Equipment	units	Retrofit Case
Steam Turbine Generator	kW	463,044
Flue Gas Expander Generator	kW	21,018
Recycle Gas Expander Generator	kW	2,834
Gross Power Generation	kW	486,896
Existing Plant Auxiliary Loads	kW	33,430
CO_2 Capture Process Auxiliary Loads	kW	142,924
Net Power Generation	kW	310,542

Individual auxiliary loads for the newly installed equipment are presented in Exhibit 4-5.

Exhibit 4-5
Retrofitted Unit 5 Auxiliary Loads

Equipment	Units	Retrofit Case
Booster Air Fan	kW	3,721
Flue Gas Compressors	kW	54,610
Vacuum Pumps	kW	24,063
CO_2 Compressors	kW	42,649
CO_2 Dryer	kW	133
Chiller Compressors	kW	13,009
CO_2 Pump	kW	2,282
Auxiliary Cooling Service	kW	2,457
Total Auxiliary Load	kW	142,924

All of the equipment required for the efficient operation of the MTR CO_2 capture process is not currently commercially available. Turbomachinery vendors who were contacted for this study indicated that while there may not currently be off-the-shelf equipment for this particular application the potential exists to engineer the required equipment from industry standard designs. Vendors also indicated a willingness to develop the equipment should the market develop. As a result, the study assumes the future achievable performance characteristics for such pieces of equipment.

4.5.2 Water Balance

The addition of CO_2 capture and compression at the Conesville station will require that additional heat be rejected to the atmosphere by the plant. A new cooling tower is designed to provide cooling water service to the new equipment. Additional make-up water will be required by the plant in order to operate. Water is recovered to offset the increase in demand through the additional cooling of the flue gas in the DCC and as a result of water permeating through the membranes with the CO_2. Blowdown from the plant back to the water source will also increase. Four cycles of concentration are assumed based on the current operation of the existing cooling system. Nominal values for the increases in these flows are presented below in Exhibit 4-6. This assessment assumes that the only changes to the plants' water balance are a result of the heat rejected through the new cooling tower.

Exhibit 4-6
Incremental Water Balance

	Units	Base Case	MTR-1
Total Cooling Tower Make-up	gpm	5,014	7,327
Difference in CT MU	gpm	Base	2,313
Water Recovered	gpm	NA	788
Net Change in Makeup due to Retrofit	**gpm**	**Base**	**1,525**
Cooling Tower Blowdown	gpm	1,253	1,832
Net Change in Blowdown due to Retrofit	**gpm**	**Base**	**578**

4.5.3 Flue Gas Analysis and Emissions

As a result of the retrofitted CO_2 capture system, the flue gas which is discharged will change. The temperature and pressure of the flue gas entering the stack have been held constant to allow for the necessary buoyancy to carry the gas out of the stack and ensure sufficient mixing in the atmosphere. The changes in the composition of the flue gas are presented in Exhibit 4-7.

Exhibit 4-7
Stack Gas Composition

Constituent	units	Base Case	Retrofit Case
Argon	mol %	0.80	1.08
Carbon Dioxide	mol %	12.56	1.69
Water	mol %	16.36	1.24
Nitrogen	mol %	67.08	90.93
Nitrous Oxides	mol %	0.02	0.03
Oxygen	mol %	3.18	5.03
Sulfur Dioxide	mol %	0.00 (ca 50 ppm)	0.00 (ca 2 ppm)

In addition to capturing CO_2 the membranes also reduce the stack emissions of NO_x and SO_2 which permeate through membrane because they are polar molecules. Post retrofit reduction in NOx and SO_2 emissions is illustrated in Exhibit 4-8.

Exhibit 4-8
NOx and SOx Emissions

Constituent	units	Base Case	Retrofit Case
Nitrous Oxides	tons/day	15.1	11.7
Sulfur Dioxide	tons/day	6.3	0.2

The primary goal of the membranes is to capture CO_2. Sufficient membrane area is installed so that nominally 90% of the CO_2 is captured. Reduction in the rate of CO_2 emissions is presented in Exhibit 4-9.

Exhibit 4-9
CO_2 Emissions

Constituent	units	Base Case	Retrofit Case
Carbon Dioxide	lbs/hr	866,102	90,007
Carbon Dioxide	tons/day	10,393	1,080

4.5.4 Product CO_2 Quality

The product CO_2 purity achieved by the membrane and purification systems meets or exceeds the specified CO_2 quality as presented in Exhibit 4-10.

Exhibit 4-10
Product CO_2 Composition.

Parameter	Units	Specified Value	Retrofit Case
Pressure	psia	2,015	2,015
CO_2, min	vol%	96%	99.98%
H_2O, max	vol%	0.015%	0.000%
N_2, max	vol%	0.6%	0.01%
O_2, max	ppmv	100	100
SO_2, max	vol%	1%	0% (< 1 ppm)

Note: The composition values listed for the retrofit case are mol%.

4.6 OPTIMIZATION OF CO_2 SYSTEM INTEGRATION WITH THE EXISITNG PLANT:

Two equipment arrangements for the CO_2 compression system were evaluated. The compressors were staged to achieve gas temperatures on the order of 300°F in the first arrangement. This elevated gas temperature allows for the recovery of the heat generated by the compression. Compression stages are cooled more efficiently in the second case by utilizing circulating cooling water. The purpose of evaluating these two cases was to provide the most cost and energy efficient system within the constraints of the existing condenser and steam turbine.

An increase in gross generation is the target of the first arrangement. Cold condensate is routed to the intercoolers in the CO_2 compression area. The condensate serves as the heat sink for the heat generated in compressing the CO_2. Heated condensate is sent back to the steam cycle's deaerator. Low pressure extractions from the steam turbine are thus reduced. Modeling of the steam cycle showed that this arrangement resulted in an increased steam flow that could not be processed by the existing turbine due to exhaust end limitations. An auxiliary steam turbine generator is installed to generate additional power. This allowed all of the available steam flow to be utilized. A new condenser is installed for the auxiliary steam turbine to hold down the back pressure of both turbines.

The second arrangement replaces the condensate as the cooling medium with circulating water. Colder interstage compression temperatures can be achieved with the circulating water. Additionally the gas can be cooled more frequently since it is not necessary to have the gas temperature rise to 300°F. A more efficient compressor with a lower power demand results from this design.

Net power output from the facility was comparable between the two arrangements. All of the newly installed compression and cooling systems in the second arrangement operate independently of the existing facility. In the first arrangement, a new steam turbine, condenser, piping and ancillary pumps and valves are required in addition to the compression system. The steam turbine and condenser would tie in at multiple points to the existing steam cycle and require all of the controls necessary for an operable system. The second arrangement was selected over the first as a result of the projected cost and complexity of the first arrangement.

5. COST ANALYSIS

This section presents the cost estimate basis, cost methodology, capital and O&M costs for the MTR case, as well as the escalated costs for the MEA-1 case.

5.1 COST ESTIMATE BASIS

The basis for capital and operating costs estimates in this study will be consistent with the basis in the 2007 DOE/NETL Carbon Dioxide Capture from Existing Coal-Fired plants study [1], except that the costs analysis in this report will be expressed in December 2009 U.S. dollars. This approach will enable comparison of this study results with the appropriately escalated results of the 2007 DOE/NETL study [1].

The capital cost estimates in this study will be assessed on a Total Investment Cost (TIC) level[6], and will be presented on an engineering, procurement, and construction (EPC) reimbursable basis with process and project contingencies. All costs will be estimated in December 2009 U.S. dollars. These costs will include all required equipment to complete the retrofit such as the new membrane based CO_2 capture system, the new CO_2 compression and dehydration system, additional new balance of plant systems, and modifications to the existing plant equipment and systems as required to support operation of the retrofitted plant.

Operating and maintenance (O&M) costs will be calculated for all systems. The O&M costs for the Base Case (pre-retrofit Conesville #5 Unit) will be based on the 2007 DOE/NETL study [1]. For the retrofit CO_2 capture system evaluations, additional O&M costs will be calculated for the new equipment. The variable operating and maintenance (VO&M) costs for the new equipment include such categories as chemicals, membrane replacement, maintenance material and labor, and make-up power cost (MUPC) from the reduction in net electricity production. The fixed operating and maintenance (FO&M) costs for the new equipment includes operating labor and maintenance.

The following assumptions were made in developing cost estimates for the retrofit case:

1. December 2009 U.S. dollars
2. Outdoor installation, except for the equipment[7] in the MTR CO_2 capture system area, which will be indoors
3. Investment in new utility systems (outside of the plant boundary) is outside the scope
4. CO_2 product pipeline is outside the scope
5. No special limitations for transportation of large equipment
6. No underground interferences
7. Scope includes costs for treating incremental makeup water

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[6] The TIC cost documented in the reference study is consistent with the Total Plant Cost (TPC) nomenclature utilized by DOE/NETL.

[7] Equipment contained in the CO_2 capture building includes membrane modules A&B, vacuum pumps, booster fan, flue gas compressors, and direct contact cooler.

8. All waste streams can be discharged to the source, and additional waste water treatment is not necessary
9. No protection against unusual airborne contaminants (dust, salt, etc.)
10. No unusual wind storms
11. No earthquakes
12. No piling required
13. All releases can go to atmosphere – no flare provided
14. Annual operating time is 7,446 hr/yr (85% capacity factor)
15. The investment cost estimate was developed as a factored estimate based on in-house data for the major equipment, supplemented by budgetary quotes as available. Such an estimate can be expected to have accuracy of ±30%.
16. Process and project contingency will be added to the EPC to derive the TIC.
17. Make-up power cost was assessed at a 20-year levelized rate of 7.10 ¢/kWh (equivalent to a new Subcritical Pulverized Coal (Greenfield) Plant without carbon capture)
    Note: This is relevant to the EPRI Economics and not the capital cost development.
18. No purchases of utilities or charges for shutdown time have been charged against the project

Other exclusions from the cost estimate are as follows:

1. Soil investigation
2. Environmental permits
3. Disposal of hazardous or toxic waste
4. Disposal of existing materials
5. Custom's and Import duties
6. Sales/use tax
7. Forward escalation
8. Capital spare parts
9. Chemical loading facilities
10. Buildings except for MTR module building
11. Financing cost
12. Owners cost
13. Guards during construction
14. Site medical and ambulance service
15. Cost & fees of authorities

5.2 COST METHODOLOGY

This section presents a methodology for estimating incremental costs associated with the CO_2 membrane retrofit and CO_2 capture operation.

The TIC and Operation and Maintenance (O&M) costs developed herein have an accuracy level of ±30 percent, consistent with the screening study level of information available for this study. All cases were evaluated assuming base load operation and a plant capacity factor (CF) of 85% based on reference [1].

An in-house database and conceptual estimating models were used for the capital cost and O&M cost estimates. Costs were further calibrated using a combination of adjusted vendor-furnished and actual cost data from recent design and design/build projects. In addition, vendor quotes where obtained for select major equipment to improve the accuracy of the model.

All capital and O&M costs are presented as "overnight costs" expressed in December 2009 dollars.

Capital costs are presented at the TIC level. TIC includes:

- Equipment (complete with initial chemical and catalyst loadings);
- Materials;
- Labor (direct and indirect);
- Engineering and construction management;
- Contingencies (process and project);

Owner's costs are excluded.

In this study, this CO_2 retrofit case was treated as a first of a kind (FOAK) in that process contingency was applied to select equipment. Non reoccurring engineering (NRE) costs were not included. For example the NRE costs associated with the customization / development of the low temperature gas expander were excluded.

5.2.1 <u>Estimate Scope</u>

The estimate represents a retrofit CO_2 capture facility on the existing Conesville #5 Unit.

The estimate boundary limit is defined as the total retrofit plant facility within the "fence line". The battery limits for the estimate extend from the outlet of the existing FGD unit to the inlet of the existing stack, and the inlet of the booster air fans to the inlet of the existing secondary air fan.

5.2.2 <u>Membrane Cost</u>

The membrane cost was provided by MTR as presented in Exhibit 5-1. MTR has estimated the average module lifetime as 3 years for the flue gas application based on their natural gas business where the average module lifetime is more than 5 years. [28] WorleyParsons has based the 3 year life on a 100% capacity factor. Thus, based on an equivalent capacity factor of 85%, the modules would be replaced after 3.5 years of service.

Exhibit 5-1
Membrane Cost

Constituent	units	Base Case	Notes
Installed Membrane Cost	USD, Dec 09	$50/m2	Includes module housings
Replacement Module Cost	USD, Dec 09	$10/m2	20% of installed cost

5.2.3 Construction Labor

Costs for construction labor are based on published union labor rates for the greater Columbus, Ohio area. The estimate reflects a 5-day, 10 hour per day work week. No further provisions for attracting or retaining craft labor were included. The estimate assumes that an adequate and qualified labor pool will be available to staff the project.

5.2.4 Contingency

Consistent with the NETL quality guidelines [29], both process contingency and project contingency have been applied to this cost estimate. **"Process contingency** is designed to compensate for uncertainty in cost estimates caused by **performance uncertainties** associated with the **development status** of a technology." This is distinct from the **project contingency** associated with uncertainty in the cost estimate caused by an **incomplete technical definition**.

The NETL quality guidelines cite the AACE standards presented in Exhibit 5-2 for process contingency as a function of a given area's technological maturity. This standard was employed for this study.

Exhibit 5-2
AACE Standards for Process Contingency

Technology Status	Process Contingency
New technology, little or no test data	40% +
New technology, prototype test data	20-35%
Modifications to commercial technology	5-20%
Commercial technology	0-5%

As such, WorleyParsons assigned the following process contingency for the CO_2 membranes and the CO_2 purification.

1. **Membrane: Process Contingency taken as 20% of Bare Erected Cost (BEC)**
   (Basis: Membrane status judged between: "New technology, prototype data" [20-35%] and "modifications to commercial technology" [5-20%])

2. **CO2 Purification Process Contingency taken as 15% of BEC**
   (Basis: The status is judged to be consistent with "modifications to commercial

technology". [5-20%] The 15% level was chosen as reasonable since the ASPEN purification model was never calibrated to a commercial system.)

3. No process contingency (i.e., only project contingency is applied) is assumed for the following equipment:
    a. Vacuum Pump/ Compressor
    b. CO_2 Compressors
    c. Flue Gas compressors
    d. Flue Gas Expander
    e. Cryogenic Expander

    Although these compressors and expanders may need additional engineering for the first of a kind unit for this novel application, the NRE costs have not been included in the cost estimate.

This methodology differs somewhat from that employed in the reference study and presented in Exhibit 5-3. However, since the technologies differ between the MEA and the membrane retrofit applications, it is logical that the process contingencies would differ. Therefore WorleyParsons utilized the NETL guideline approach, instead of directly using the process contingency utilized in the MEA reference study.

Exhibit 5-3
Project and Process Contingencies (Used in Ref Study)

Capital Equipment	Project Contingency*	Process Contingency*
CO_2 Separation and Compression system	25%	18%
Flue gas Desulfurization (FGD)	11%	0%

* Percent of bare erected cost (i.e. subtotal direct cost in the investment tables for each case)

5.2.5 Retrofit Cost Considerations

A retrofit project will typically have labor costs that are higher than a new or greenfield project because of the congestion and construction difficulty associated with implementation around existing equipment and possibly an operating plant.

To maximize compatibility with the reference amine study [1] it would be desirable to utilize the same retrofit labor cost factor between the studies. Unfortunately, the retrofit cost factor was not specified in the subject report. It is suspected that the reference study did not account for this retrofit labor cost premium.

In this report, a 10% retrofit cost factor was added to labor costs.

5.2.6 Exclusions

As discussed above, the scope of the estimates is generally limited to scope within the project fence. The following items are excluded from the estimate:

- Escalation to period of performance
- Owner's Costs (see below)

- Premiums beyond 5-10's and per diem required to attract craft labor
- Craft Bussing
- Permits
- Warranty
- Builder's all-Risk insurance
- Premiums associated with an EPC contracting approach
  Including, but not limited to, Performance guarantees, LD's, etc.
- Allowance for Funds Used During Construction
- All taxes with the exception of payroll taxes
- Disposal costs for site-borne contaminated or hazardous materials
- Costs for unknown underground obstructions or interferences
- Cost for electricity consumed during startup
- Costs for fuel consumed during startup
- Costs for water consumed during startup
- Continuous Emissions Monitoring Equipment & Bulk Materials
- Demolition
- Relocation of any existing utilities (underground or aboveground)
- Environmental remediation
- Piles (if required)

Typical Owner's Costs

Typical owner's costs include, but are not limited to:

- Permits & Licensing (other than construction permits)
- Land Acquisition / Rights of Way Costs
- Economic Development
- Project Development Costs
- Environmental Impact Costs
- Excessive Noise Abatement
- Local Facilitation Costs
- Improvement to existing roads or infrastructure
- Legal Fees
- Wetland Mitigation
- Interconnection Agreements
- Fuel Purchase Agreements
- Owner's Engineering / Project & Construction Management Staff
- Plant Operators during startup
- Electricity consumed during startup
- Fuel and Reagent consumed during startup
- Initial Fuel & Reagent Inventory
- Operating Spare Parts
- Mobile Equipment for use during plant operations
- Furnishings for new Office, Warehouse and Laboratory
- Financing Costs
- Owner's Contingency

5.2.7 Maintenance Material and Labor Costs

The maintenance material and labor costs were evaluated on the basis of relationships of maintenance cost to initial capital cost. The values presented in the O&M summary represent a weighted analysis in which the individual cost relationships were considered for each major plant component or section.

5.3 CAPITAL AND O&M COST RESULTS

The capital cost summary for the MTR retrofit case is presented in Exhibit 5-4. Additional cost details for this plant are presented in Exhibit 5-5.

Exhibit 5-4
Total Plant Cost for MTR-1: Cost Summary

Client: DOE/NETL/BAH — Report Date: 2010-Dec-23
Project: MTR CO2 Membrane for Capturing CO2 from Power Plant Flue Gas

TOTAL PLANT COST SUMMARY

Case: MTR CO2 Membrane Retrofit
Plant Size: 391.2 TPH CO2 — Estimate Type: Conceptual — Cost Base (Dec) 2009 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $ x 1000	TOTAL PLANT COST $K / TPH
1	COAL & SORBENT HANDLING	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
2	COAL & SORBENT PREP & FEED	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
3	FEEDWATER & MISC. BOP SYSTEMS	$3,657	$0	$1,910	$0	$0	**$5,567**	$516	$0	$1,217	**$7,300**	$19
4	PC BOILER & ACCESSORIES	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
5A	FLUE GAS TREATMENT	$64,173	$25,106	$51,271	$0	$0	**$140,550**	$13,542	$0	$30,473	**$184,565**	$472
5B	CO2 PURIFICATION & COMPRESSION	$122,359	$10,891	$57,103	$0	$0	**$190,353**	$17,893	$8,400	$43,329	**$259,974**	$665
6	COMBUSTION TURBINE/ACCESSORIES	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
7	HRSG, DUCTING & STACK	$6,378	$1,959	$10,121	$0	$0	**$18,457**	$1,647	$0	$3,211	**$23,315**	$60
8	STEAM TURBINE GENERATOR	$18	$0	$70	$0	$0	**$88**	$8	$0	$19	**$115**	$0
9	COOLING WATER SYSTEM	$6,508	$2,965	$6,492	$0	$0	**$15,966**	$1,480	$0	$2,462	**$19,908**	$51
10	ASH/SPENT SORBENT HANDLING SYS	$0	$0	$0	$0	$0	**$0**	$0	$0	$0	**$0**	$0
11	ACCESSORY ELECTRIC PLANT	$9,747	$6,166	$26,205	$0	$0	**$42,118**	$3,896	$0	$6,080	**$52,094**	$133
12	INSTRUMENTATION & CONTROL	$5,887	$0	$9,153	$0	$0	**$15,040**	$1,405	$0	$2,041	**$18,487**	$47
13	IMPROVEMENTS TO SITE	$1,953	$1,123	$5,746	$0	$0	**$8,823**	$866	$0	$1,938	**$11,627**	$30
14	BUILDINGS & STRUCTURES	$6,364	$505	$6,758	$0	$0	**$13,627**	$1,242	$0	$2,230	**$17,100**	$44
	TOTAL COST	**$227,044**	**$48,715**	**$174,829**	**$0**	**$0**	**$450,588**	**$42,496**	**$8,400**	**$93,000**	**$594,484**	$1,520

Exhibit 5-5
Total Plant Cost for MTR-1: Cost Details

Client: DOE/NETL/BAH — **Report Date:** 2010-Dec-23
Project: MTR CO2 Membrane for Capturing CO2 from Power Plant Flue Gas

TOTAL PLANT COST SUMMARY

Case: MTR CO2 Membrane Retrofit
Plant Size: 391.2 TPH CO2 — **Estimate Type:** Conceptual — **Cost Base (Dec)** 2009 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $	TOTAL PLANT COST $K / TPH
1	COAL & SORBENT HANDLING											
1.1	Coal Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.2	Coal Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.3	Coal Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.4	Other Coal Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.5	Sorbent Receive & Unload	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.6	Sorbent Stackout & Reclaim	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.7	Sorbent Conveyors	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.8	Other Sorbent Handling	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
1.9	Coal & Sorbent Hnd.Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 1.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
2	COAL & SORBENT PREP & FEED											
2.1	Coal Crushing & Drying	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.2	Coal Conveyor to Storage	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.3	Coal Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.4	Misc.Coal Prep & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.5	Sorbent Prep Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.6	Sorbent Storage & Feed	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.7	Sorbent Injection System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.8	Booster Air Supply System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
2.9	Coal & Sorbent Feed Foundation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 2.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
3	FEEDWATER & MISC. BOP SYSTEMS											
3.1	Feedwater System	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.2	Water Makeup & Pretreating	$3,056	$0	$1,435	$0	$0	$4,490	$416	$0	$981	$5,888	$15
3.3	Other Feedwater Subsystems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.4	Service Water Systems	$601	$0	$475	$0	$0	$1,077	$100	$0	$235	$1,412	$4
3.5	Other Boiler Plant Systems	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.6	FO Supply Sys & Nat Gas	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.7	Waste Treatment Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
3.8	Misc. Equip.(cranes,AirComp.,Comm.)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 3.	**$3,657**	**$0**	**$1,910**	**$0**	**$0**	**$5,567**	**$516**	**$0**	**$1,217**	**$7,300**	**$19**
4	PC BOILER & ACCESSORIES											
4.1	PC Boiler, BHse & Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.2	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.3	Open	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.4	Boiler BoP (w/ ID Fans)	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
4.5	Primary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.6	Secondary Air System	w/4.1	$0	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.8	Major Component Rigging	$0	w/4.1	w/4.1	$0	$0	$0	$0	$0	$0	$0	$0
4.9	Boiler Foundations	$0	w/14.1	w/14.1	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 4.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
5A	FLUE GAS TREATMENT											
5A.1	In-situ Flue Gas / Glycol Heat HX	$2,031	$880	$1,760	$0	$0	$4,671	$450	$0	$1,024	$6,145	$16
5A.2	Flue Gas Compressor	$41,000	$16,400	$32,800	$0	$0	$90,200	$8,696	$0	$19,779	$118,675	$303
5A.3	Direct Contact Cooler	$1,855	$773	$1,546	$0	$0	$4,174	$402	$0	$915	$5,492	$14
5A.4	In-situ Glycol / Flue Gas Heat HX (HX1-C)	$3,066	$1,329	$2,657	$0	$0	$7,052	$680	$0	$1,546	$9,278	$24
5A.5	Glycol Circulation Pumps	$124	$124	$248	$0	$0	$497	$48	$0	$109	$654	$2
5A.6	Flue Gas Turbo Expander	$14,000	$5,600	$11,200	$0	$0	$30,800	$2,969	$0	$6,754	$40,523	$104
5A.9	Booster Air Fan	$2,097	$0	$1,059	$0	$0	$3,156	$296	$0	$345	$3,797	$10
	SUBTOTAL 5.	**$64,173**	**$25,106**	**$51,271**	**$0**	**$0**	**$140,550**	**$13,542**	**$0**	**$30,473**	**$184,565**	**$472**
5B	CO2 PURIFICATION & COMPRESSION											
5B.1	MTR CO2 Membrane Modules	$13,657	$0	$13,657	$0	$0	$27,314	$2,612	$5,463	$7,078	$42,467	$109
5B.2	Compression Systems	$77,957	$2,307	$26,257	$0	$0	$106,521	$9,875	$0	$23,279	$139,675	$357
5B.3	CO2 Purification Systems	$13,765	$1,931	$3,884	$0	$0	$19,580	$1,873	$2,937	$4,878	$29,268	$75
5B.4	CO2 Chilling Systems	$16,981	$6,652	$13,305	$0	$0	$36,938	$3,533	$0	$8,094	$48,565	$124
	SUBTOTAL 5B.	**$122,359**	**$10,891**	**$57,103**	**$0**	**$0**	**$190,353**	**$17,893**	**$8,400**	**$43,329**	**$259,974**	**$665**
6	COMBUSTION TURBINE/ACCESSORIES											
6.1	Combustion Turbine Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.2	Combustion Turbine Accessories	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
6.3	Compressed Air Piping	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
6.9	Combustion Turbine Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 6.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
7	HRSG, DUCTING & STACK											
7.1	Heat Recovery Steam Generator	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
7.2	HRSG Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.3	Ductwork	$6,378	$0	$8,518	$0	$0	$14,896	$1,306	$0	$2,430	$18,632	$48
7.4	Stack	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
7.9	Duct & Stack Foundations	$0	$1,959	$1,603	$0	$0	$3,562	$341	$0	$780	$4,683	$12
	SUBTOTAL 7.	**$6,378**	**$1,959**	**$10,121**	**$0**	**$0**	**$18,457**	**$1,647**	**$0**	**$3,211**	**$23,315**	**$60**
8	STEAM TURBINE GENERATOR											
8.1	Steam TG & Accessories	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.2	Turbine Plant Auxiliaries	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.3	Condenser & Auxiliaries	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
8.4	Steam Piping	$18	$0	$70	$0	$0	$88	$8	$0	$19	$115	$0
8.9	TG Foundations	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 8.	**$18**	**$0**	**$70**	**$0**	**$0**	**$88**	**$8**	**$0**	**$19**	**$115**	**$0**

Exhibit 5-5
Total Plant Cost for MTR-1: Cost Details (Cont'd)

Client: DOE/NETL/BAH **Report Date:** 2010-Dec-23
Project: MTR CO2 Membrane for Capturing CO2 from Power Plant Flue Gas

TOTAL PLANT COST SUMMARY

Case: MTR CO2 Membrane Retrofit
Plant Size: 391.2 TPH CO2 **Estimate Type:** Conceptual **Cost Base (Dec)** 2009 ($x1000)

Acct No.	Item/Description	Equipment Cost	Material Cost	Labor Direct	Labor Indirect	Sales Tax	Bare Erected Cost $	Eng'g CM H.O.& Fee	Contingencies Process	Contingencies Project	TOTAL PLANT COST $	TOTAL PLANT COST $K / TPH
9	COOLING WATER SYSTEM											
9.1	Cooling Towers	$4,601	$0	$371	$0	$0	$4,972	$453	$0	$543	$5,968	$15
9.2	Circulating Water Pumps	$1,213	$0	$63	$0	$0	$1,276	$105	$0	$138	$1,519	$4
9.3	Circ.Water System Auxiliaries	$130	$0	$25	$0	$0	$155	$14	$0	$17	$186	$0
9.4	Circ.Water Piping	$0	$1,931	$2,706	$0	$0	$4,637	$433	$0	$761	$5,831	$15
9.5	Make-up Water System	$211	$0	$410	$0	$0	$621	$60	$0	$102	$783	$2
9.6	Component Cooling Water Sys	$353	$0	$408	$0	$0	$762	$72	$0	$125	$958	$2
9.9	Circ.Water System Foundations& Structures	$0	$1,035	$2,508	$0	$0	$3,543	$343	$0	$777	$4,662	$12
	SUBTOTAL 9.	**$6,508**	**$2,965**	**$6,492**	**$0**	**$0**	**$15,966**	**$1,480**	**$0**	**$2,462**	**$19,908**	**$51**
10	ASH/SPENT SORBENT HANDLING SYS											
10.1	Ash Coolers	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.2	Cyclone Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.3	HGCU Ash Letdown	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.4	High Temperature Ash Piping	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.5	Other Ash Recovery Equipment	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
10.6	Ash Storage Silos	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.7	Ash Transport & Feed Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.8	Misc. Ash Handling Equipment	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
10.9	Ash/Spent Sorbent Foundation	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 10.	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**	**$0**
11	ACCESSORY ELECTRIC PLANT											
11.1	Generator Equipment	$425	$0	$100	$0	$0	$525	$47	$0	$43	$615	$2
11.2	Station Service Equipment	$2,906	$0	$1,460	$0	$0	$4,366	$414	$0	$358	$5,138	$13
11.3	Switchgear & Motor Control	$3,367	$0	$869	$0	$0	$4,235	$394	$0	$463	$5,092	$13
11.4	Conduit & Cable Tray	$0	$2,215	$11,078	$0	$0	$13,293	$1,300	$0	$2,189	$16,782	$43
11.5	Wire & Cable	$0	$3,884	$11,671	$0	$0	$15,554	$1,393	$0	$2,542	$19,489	$50
11.6	Protective Equipment	$136	$0	$707	$0	$0	$843	$84	$0	$93	$1,019	$3
11.7	Standby Equipment	$374	$0	$13	$0	$0	$387	$36	$0	$42	$465	$1
11.8	Main Power Transformers	$2,540	$0	$57	$0	$0	$2,597	$199	$0	$280	$3,075	$8
11.9	Electrical Foundations	$0	$67	$251	$0	$0	$318	$31	$0	$70	$419	$1
	SUBTOTAL 11.	**$9,747**	**$6,166**	**$26,205**	**$0**	**$0**	**$42,118**	**$3,896**	**$0**	**$6,080**	**$52,094**	**$133**
12	INSTRUMENTATION & CONTROL											
12.1	PC Control Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.2	Combustion Turbine Control	N/A	$0	N/A	$0	$0	$0	$0	$0	$0	$0	$0
12.3	Steam Turbine Control	w/8.1	$0	w/8.1	$0	$0	$0	$0	$0	$0	$0	$0
12.4	Other Major Component Control	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
12.5	Signal Processing Equipment	w/12.7	$0	w/12.7	$0	$0	$0	$0	$0	$0	$0	$0
12.6	Control Boards,Panels & Racks	$304	$0	$278	$0	$0	$582	$56	$0	$96	$734	$2
12.7	Distributed Control System Equipment	$3,067	$0	$820	$0	$0	$3,887	$364	$0	$425	$4,676	$12
12.8	Instrument Wiring & Tubing	$1,650	$0	$5,046	$0	$0	$6,695	$603	$0	$1,095	$8,393	$21
12.9	Other I & C Equipment	$867	$0	$3,009	$0	$0	$3,875	$382	$0	$426	$4,684	$12
	SUBTOTAL 12.	**$5,887**	**$0**	**$9,153**	**$0**	**$0**	**$15,040**	**$1,405**	**$0**	**$2,041**	**$18,487**	**$47**
13	IMPROVEMENTS TO SITE											
13.1	Site Preparation	$0	$33	$958	$0	$0	$991	$100	$0	$218	$1,309	$3
13.2	Site Improvements	$0	$1,090	$1,976	$0	$0	$3,066	$301	$0	$673	$4,040	$10
13.3	Site Facilities	$1,953	$0	$2,812	$0	$0	$4,765	$466	$0	$1,046	$6,277	$16
	SUBTOTAL 13.	**$1,953**	**$1,123**	**$5,746**	**$0**	**$0**	**$8,823**	**$866**	**$0**	**$1,938**	**$11,627**	**$30**
14	BUILDINGS & STRUCTURES											
14.1	MTR CO2 Capture Building	$5,247	$0	$3,759	$0	$0	$9,006	$820	$0	$1,474	$11,300	$29
14.2	CO2 Compression & Purification Area	$1,117	$0	$2,341	$0	$0	$3,458	$316	$0	$566	$4,340	$11
14.3	Administration Building	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.4	Circulation Water Pumphouse	$0	$122	$142	$0	$0	$264	$24	$0	$43	$331	$1
14.5	Water Treatment Buildings	$0	$384	$516	$0	$0	$899	$82	$0	$147	$1,129	$3
14.6	Machine Shop	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.7	Warehouse	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.8	Other Buildings & Structures	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
14.9	Waste Treating Building & Str.	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0	$0
	SUBTOTAL 14.	**$6,364**	**$505**	**$6,758**	**$0**	**$0**	**$13,627**	**$1,242**	**$0**	**$2,230**	**$17,100**	**$44**
	TOTAL COST	**$227,044**	**$48,715**	**$174,829**	**$0**	**$0**	**$450,588**	**$42,496**	**$8,400**	**$93,000**	**$594,484**	**$1,520**

The operating and maintenance cost for the MTR-1 case is presented in Exhibit 5-6.

Exhibit 5-6
Operating & Maintenance Cost for MTR-1

INITIAL & ANNUAL O&M EXPENSES					Cost Base (Dec):	2009
MTR CO2 Membrane Retrofit					Heat Rate-net (Btu/kWh):	13,762
					CO2 TPH:	391
					Capacity Factor (%):	85
OPERATING & MAINTENANCE LABOR						
Operating Labor						
Operating Labor Rate (base):	34.65	$/hour				
Operating Labor Burden:	30.00	% of base				
Labor O-H Charge Rate:	25.00	% of labor				
			Total			
Skilled Operator	0.0		0.0			
Operator	2.0		2.0			
Foreman	0.0		0.0			
Lab Tech's, etc.	0.0		0.0			
TOTAL-O.J.'s	2.0		2.0			
					Annual Cost	Annual Unit Cost
					$	$/TPH
Annual Operating Labor Cost					$789,188	$2.018
Maintenance Labor Cost					$2,956,092	$7.557
Administrative & Support Labor					$936,320	$2.394
TOTAL FIXED OPERATING COSTS					**$4,681,601**	**$11.969**
VARIABLE OPERATING COSTS						
						$/TPH
Maintenance Material Cost					**$4,434,138**	**$0.00152**
Consumables	Consumption Initial	Consumption /Day	Unit Cost	Initial Cost		
Water (/1000 gallons)	0	3,744.00	1.23	**$0**	**$1,428,423**	**$0.00049**
Chemicals						
MU & WT Chem.(lbs)	63,432	9,061.68	0.20	$12,481	$553,158	$0.00019
Replacement Membrane Modules (m2)	0	498.88	10.00	$0	$1,547,779	$0.00053
Subtotal Chemicals				**$12,481**	**$2,100,938**	**$0.00072**
Other						
Supplemental Fuel (MBtu)	0	0	0.00	$0	$0	$0.00000
Supplemental Electricity (for consumption) (MWh)	0	2,958	71.00	$0	$65,150,683	$0.02237
Subtotal Other				**$0**	**$65,150,683**	**$0.02237**
Waste Disposal						
Fly Ash (ton)	0	0.00	18.45	$0	$0	$0.00000
Bottom Ash (ton)	0	0.00	18.45	$0	$0	$0.00000
Subtotal-Waste Disposal				**$0**	**$0**	**$0.00000**
By-products & Emissions						
CO2 Product (ton)	0	9,388	0.00	$0	$0	$0.00000
Subtotal By-Products				**$0**	**$0**	**$0.00000**
TOTAL VARIABLE OPERATING COSTS				**$12,481**	**$73,114,182**	**$0.02510**
Fuel (ton)	0	0	0.00	**$0**	**$0**	**$0.00000**

5.4 COST SENSITIVITIES

Several sensitivities were performed and are documented herein.

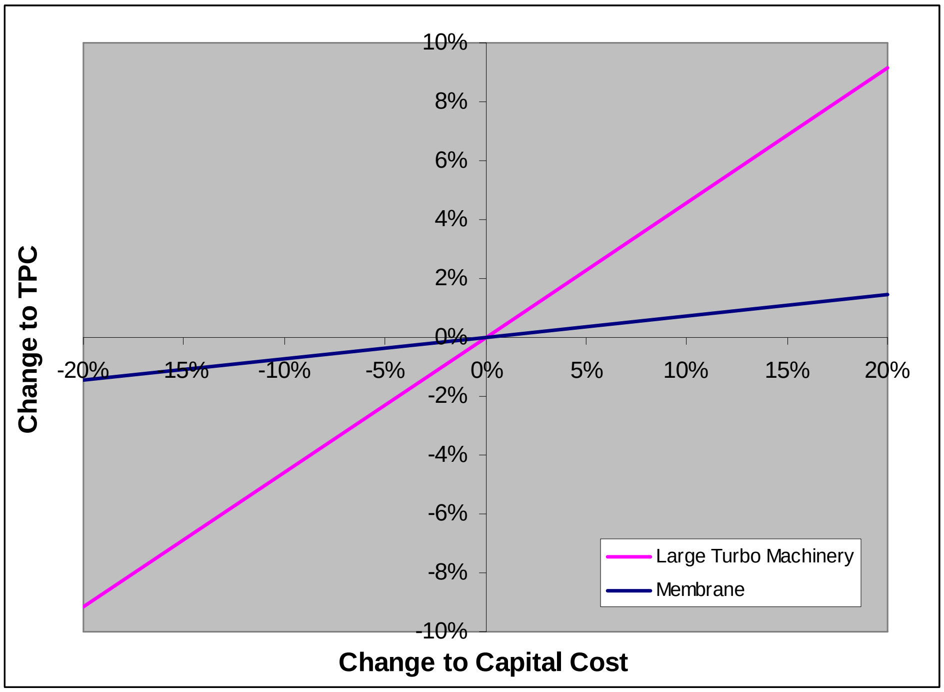
5.4.1 Sensitivities of Select Cost Area

Sensitivities were performed on a few select cost areas in order to understand their impact on the retrofit TPC as shown in Exhibit 5-7 and Exhibit 5-8.

Exhibit 5-7
Table of Select Cost Sensitivities

Sensitivity Area	Change	TPC Low Value	TPC High Value	TPC Range
Membrane Cost	±20%	$585,990,000	$602,978,000	±1.4%
Large Turbo Machinery	±20%	$540,196,000	$648,880,000	±9.2%

Exhibit 5-8
Plot of Select Cost Sensitivities



Per the sensitivity plot above, it can be clearly seen that the retrofit cost is not overly sensitivity to a 20% variations in the membrane cost. In contrast, a 20% variation in the large turbo machinery[8] cost clearly has a much greater impact on the retrofit cost than the membrane. As such it may be advantageous to evaluate configurations that could minimize or eliminate the turbo machinery.

5.4.2 Sensitivity Reflecting an Advanced Membrane

A high-level cost sensitivity was performed to evaluate what would happen to the retrofit cost if advanced membranes were utilized to reduce the membrane operating pressure from 2.0 bar to 1.2 bar. MTR notes that "With advanced membranes, it is possible to achieve the same stream conditions leaving membrane units A and B using 1.2 bar feed instead of 2.0 bar." [30]

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8 The turbo machinery included in this sensitivity are the flue gas compressors, vacuum pumps, CO_2 compressors, and flue gas expanders.

With a membrane feed pressure of only 1.2 bar, the new process would omit the FG expander as the remaining pressure ratio will not merit expanding. The major process and cost changes would therefore include the following.

1. Replacement of the Flue Gas Compressor with a FG Booster Fan
2. Removal of the Flue Gas Expander
3. Removal of the Flue Gas Heat Exchanger which transferrs heat from the flue gas going to the FGD to the flue gas ahead of the former FG Expander
4. Miscellaneous changes (heat exchange duty, etc.)

These changes yield an estimated TPC of approximately $426.0 million, or a decrease of approximately $168 million (-28%). This assumes the membrane cost would not increase.

Elimination of the above equipment also facilitates a decrease in the auxiliary load. The corresponding net power is estimated to be 328,946 kW.

5.4.3 Sensitivity of a Relaxed CO_2 Product Specification (O_2 Level)

A high-level cost sensitivity was performed to evaluate what would happen to the retrofit cost if the 100 ppm O_2 specification listed for the CO_2 product in Exhibit 2-7 was removed. This is of interest since the oxygen requirement is under debate and since the membrane process requires a distillation column based purification system in order to meet the 100 ppm O_2 level. In contrast, a chemically based CO_2 separation process like an amine system is by nature very selective to the CO_2 and does not require the purification system. Since the amine based system cost would be the same regardless of the O_2 specification in the product CO_2, the 100 ppm O_2 requirement may place an unfair bias on the membrane based process when it is compared to the amine process.

A sensitivity study was performed with no requirement on the O_2 level, but while honoring all other requirement specified in Exhibit 2-7, including the 0.6% N_2 specification. The relaxed O_2 requirement allows the distillation column to be replaced by a two stage flash process which is designed to meet the N_2 specification. The resulting O_2 level in the CO_2 product is estimated as approximately 1,400 ppm. The corresponding TPC has been estimated as approximately $593.4 millions, which represents a savings of about $1.0 million.

5.5 AMINE CO_2 SCRUBBING COST ESTIMATE ESCALATION

The CO_2 capture studies presented in the November 2007 DOE/NETL report "Carbon Dioxide Capture from Existing Coal-Fired Power Plants" were used as a comparison point for this study. The cost estimate for Case 1 (MEA-1), with 90% CO_2 capture, was escalated from Jul 2006 to December 2009 to enable comparison on an even basis. The capital cost estimate presented in the reference report is shown in Exhibit 5-9, with the escalated capital costs shown in Exhibit 5-10. The O&M cost breakdown from the MEA-1 case is presented in Exhibit 5-11 and Exhibit 5-12, for the raw and escalated values respectively.

The escalation factors used were developed in-house using various sources including industry publications, vendor inputs, and cost indices.

Exhibit 5-9
MEA-1 Capital Cost Summary for 90% CO_2 Capture (Jul 2006 USD)

Acc't Code	Description	Pieces	Direct Manhours	Equipment ($1,000)	Material ($1,000)	Labor ($1,000)	Total ($1,000)	%
11000	Heaters							0.00%
11200	Exchangers & Aircoolers		25,200	19,049		466	19,515	5.17%
12000	Vessel / Filters		6,638	5,018		123	5,141	1.36%
12100	Towers / Internals		29,859	22,571		552	23,123	6.12%
12200	Reactors							0.00%
13000	Tanks							0.00%
14100	Pumps		4,431	3,350		82	3,432	0.91%
14200	Compressors		60,663	45,856		1,122	46,978	12.43%
18000	Special Equipment		5,070	3,833		94	3,927	1.04%
	Sub-Total Equipment	**140**	**131,861**	**99,677**		**2,439**	**102,116**	**27.03%**
21000	Civil		175,815		6,977	3,253	10,230	2.71%
21100	Site Preparation							0.00%
22000	Structures		46,152		4,087	854	4,941	1.31%
23000	Buildings		24,175		1,196	447	1,643	0.43%
30000	Piping		362,619		17,942	6,708	24,650	6.52%
40000	Electrical		186,804		7,974	3,456	11,430	3.03%
50000	Instruments		153,839		12,460	2,846	15,306	4.05%
61100	Insulation		131,862		5,183	2,439	7,622	2.02%
61200	Fireproofing		65,931		1,495	1,220	2,715	0.72%
61300	Painting		32,965		698	610	1,308	0.35%
	Sub-Total Commodities		**1,180,162**		**58,012**	**21,833**	**79,845**	**21.13%**
70000	Construction Indirects						35,228	9.32%
	Sub-Total Direct Cost (Bare Erected Cost)		**1,312,023**		**58,012**	**24,272**	**217,189**	**57.48%**
71000	Construction Management						2,000	0.53%
80000	Home Office Engineering						29,400	7.78%
80000	Basic Engineering						5,000	1.32%
95000	License Fee	Excluded						0.00%
19400	Vendor Reps						1,750	0.46%
19300	Spare parts						2,900	0.77%
80000	Training cost	Excluded						0.00%
80000	Commissioning	Excluded						0.00%
19200	Catalyst & Chemicals	Excluded						0.00%
97000	Freight						4,700	1.24%
96000	CGL / BAR Insurance							0.00%
91400	Escalation to July 2006 Dollars						7,200	1.91%
	Total Base Cost						**270,139**	**71.50%**
	Contractors Fee						14,300	3.78%
	Total (EPC):						**284,439**	**75.28%**
93000	Project Contingency						54,297	14.37%
93000	Process Contingency						39,094	10.35%
	Total Investment Cost (TIC):						**377,830**	**100.00%**
	FGD System Modification						22,265	
	Total Retrofit/Investment Cost						**400,095**	

Exhibit 5-10
MEA-1 Capital Cost Summary for 90% CO_2 Capture (Dec 2009 USD)

Acc't Code	Description	Pieces	Direct Manhours	Equipment ($1,000)	Material ($1,000)	Labor ($1,000)	Total ($1,000)	%
11000	Heaters							0.00%
11200	Exchangers & Aircoolers		25,200	23,145		531	23,676	5.28%
12000	Vessel / Filters		6,638	6,069		140	6,209	1.39%
12100	Towers / Internals		29,859	27,678		629	28,307	6.32%
12200	Reactors							0.00%
13000	Tanks							0.00%
14100	Pumps		4,431	4,070		93	4,164	0.93%
14200	Compressors		60,663	55,715		1,279	56,994	12.72%
18000	Special Equipment		5,070	4,700		107	4,807	1.07%
	Sub-Total Equipment	**140**	**131,861**	**121,376**		**2,779**	**124,156**	**27.71%**
21000	Civil		175,815		8,113	3,707	11,820	2.64%
21100	Site Preparation							0.00%
22000	Structures		46,152		4,955	973	5,928	1.32%
23000	Buildings		24,175		1,450	509	1,959	0.44%
30000	Piping		362,619		22,203	7,644	29,847	6.66%
40000	Electrical		186,804		9,330	3,938	13,268	2.96%
50000	Instruments		153,839		14,578	3,243	17,821	3.98%
61100	Insulation		131,862		6,027	2,779	8,806	1.97%
61200	Fireproofing		65,931		1,738	1,390	3,129	0.70%
61300	Painting		32,965		812	695	1,507	0.34%
	Sub-Total Commodities		**1,180,162**		**69,206**	**24,879**	**94,085**	**21.00%**
70000	Construction Indirects						40,142	8.96%
	Sub-Total Direct Cost (Bare Erected Cost)		**1,312,023**		**69,206**	**27,658**	**258,382**	**57.67%**
71000	Construction Management						2,279	0.51%
80000	Home Office Engineering						34,211	7.64%
80000	Basic Engineering						5,818	1.30%
95000	License Fee	Excluded						0.00%
19400	Vendor Reps						2,036	0.45%
19300	Spare parts						3,393	0.76%
80000	Training cost	Excluded						0.00%
80000	Commissioning	Excluded						0.00%
19200	Catalyst & Chemicals	Excluded						0.00%
97000	Freight						5,680	1.27%
96000	CGL / BAR Insurance							0.00%
91400	Escalation to July 2006 Dollars						8,526	1.90%
	Total Base Cost						**320,326**	**71.50%**
	Contractors Fee						16,957	3.78%
	Total (EPC):						**337,282**	**75.28%**
93000	Project Contingency						64,384	14.37%
93000	Process Contingency						46,357	10.35%
	Total Investment Cost (TIC):						**448,023**	**100.00%**
	FGD System Modification						26,916	
	TIC w/ FGD Modification						**474,940**	

Exhibit 5-11
MEA-1 O&M Cost Summary for 90% CO_2 Capture (Jul 2006 USD)

Case 1 CO2 Separation and Compression System	July 2006	
Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		2,494
Operating Labor	2,494	
Variable O&M Costs		17,644
Chemicals	10,161	
Waste Handling & Contracted Services	767	
Maintenance (Materials and Labor)	6,716	
Feedstock O&M Costs		653
Natural Gas	653	
Levelized, Make-up Power Cost		62,194
Levelized, Make-up Power Cost (@ 6.40 ¢/kWh)	62,194	

Exhibit 5-12
MEA-1 O&M Cost Summary for 90% CO_2 Capture (Dec 2009 USD)

Case 1 CO2 Separation and Compression System	December 2009	
Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		2,647
Operating Labor	2,647	
Variable O&M Costs		20,631
Chemicals	11,778	
Waste Handling & Contracted Services	889	
Maintenance (Materials and Labor)	7,964	
Feedstock O&M Costs		575
Natural Gas	575	
Levelized, Make-up Power Cost		68,996
Levelized, Make-up Power Cost (@ 7.10 ¢/kWh)	68,996	

6. RESULTS COMPARISON (BASE, MEMBRANE, AND AMINE)

Key technical performance parameters for the MTR CO_2 membrane retrofit are compared to the base case (status quo), and to the MEA-1 (SOA 2006) and MEA-1A (Advanced MEA Technology) cases in Exhibit 6-1.

Exhibit 6-1
Summary of Technical Performance for Retrofitting Conesville Unit 5

Parameter	Units	Case-0 Base Case	MTR-1 SOA 2010	MEA-1 SOA 2006	MEA-1A Advanced
Boiler Parameters					
Main Steam Flow	lbm/hr	3,131,619	3,131,600	3,131,651	3,131,651
Reheat Steam Flow (to IP turbine)	lbm/hr	2,853,607	2,851,724	2,848,739	2,848,725
Main Steam Pressure	psia	2,535	2,535	2,535	2,535
Main Steam Temp	Deg F	1,000	1,000	1,000	1,000
Reheat Steam Temp	Deg F	1,000	1,000	1,000	1,000
Boiler Efficiency	percent	88.13	87.33	88.13	88.13
Flue Gas Flow leaving Economizer	lbm/hr	4,014,743	4,795,700	4,014,743	4,014,743
Flue Gas Temperature leaving Air Heater	Deg F	311	322	311	311
Coal Heat Input (HHV)	MMBtu/hr	4,228.7	4,273.6	4,228.7	4,228.7
Coal Heat Input (HHV)	lbm/hr	374,453	378,425	374,453	374,453
CO2 Removal System Steam & Related Parameters					
Solvent Regeneration Energy	Btu/lbm-CO2		**NA**	**1,550**	**1,200**
CO2 Removal System Steam Pressure	psia	---	203	47	47
CO2 Removal System Steam Temp	Deg F	---	718	424	424
CO2 Removal System Steam Extraction Flow	lbm/hr	---	5,696	1,210,043	975,152
CO2 Removal System Heat to Cooling Tower	MMBtu/hr	-	1,075	890.2	698.2
Natural Gas Heat Input	MMBtu/hr	-	-	13	13
CO2 produced from Natural Gas usage	lbm/hr	-	-	1,492	1,492
Generation & Auxiliary Load					
Existing Steam Turbine Generator Output	kW	463,478	463,044	342,693	367,859
CO2 Removal System Turbine Generator Output	kW	-	23,852	45,321	36,083
Total Turbine Generator Output	kW	463,478	486,896	388,014	403,942
Auxilliary Power: Existing Plant	kW	29,700	33,430	29,765	29,817
Auxilliary Power: CO2 Removal System	kW	-	142,924	54,939	54,845
Net Plant Output	**kW**	**433,778**	**310,542**	**303,310**	**319,280**
Plant Performance Parameters					
Net Plant Heat Rate (HHV)	Btu/kWh	9,749	13,762	13,985	13,285
Net Plant Efficiency (HHV)	%	35.01%	24.80%	24.41%	25.69%
Energy Penalty, (percentage points of NP Eff.)	%	Base	10.21%	10.60%	9.32%
Capacity Factor	%	85%	85%	85%	85%
Plant CO2 Emissions					
Carbon Dioxide Produced	lbm/hr	866,102	872,189	867,595	867,595
Carbon Dioxide Recovered	lbm/hr	-	782,177	779,775	779,775
Carbon Dioxide Emissions	lbm/hr	866,102	90,012	87,820	87,820
Carbon Dioxide Recovered (% of Produced)	%	0.00%	89.68%	89.88%	89.88%
Specific Carbon Dioxide Emissions	lbm/kWh	1.997	0.290	0.290	0.275
Normalized Specific CO2 Emissions (Relative to Base Case)	fraction	1.000	0.145	0.145	0.138
Avoided Carbon Dioxide Emissions (as compared to Base)	lbm/kWh	---	1.707	1.707	1.722

Note: Reference [1] page 152, Table 5-2 is the source of values for Base Case, MEA-1, and MEA-1A.

Although the table above speaks for itself, several of the more significant differences will be discussed below.

In the boiler parameters section, it is clear that the main steam (MS), and reheat (RH) steam flow rate are nearly identical in all cases. The MS and RH temperatures are controlled to 1000°F in all cases. The boiler efficiency of the MTR-1 case is 0.8% lower than the other comparison cases. The boiler efficiency of the MEA cases are unaffected compared to the base case, while the MTR-1 case has a lower efficiency as a result of the increased N_2 and CO_2 traveling through the boiler yielding lower temperatures in the furnace area and a higher flue gas temperature exiting the air heater. The decreased boiler efficiency yields a coal flow increase of about 1%.

A comparison of the CO_2 removal systems steam parameters shows that the MEA cases use a very large amount (1.0 to 1.2 million lb/h) of low pressure steam for solvent regeneration, compared to a small amount (about 6,000 lb/h) of low pressure steam for the MTR case for CO_2 product drying. The MEA cases utilized natural gas for CO_2 drying. The heat rejection for the MTR case is higher than the MEA cases as it represents heat not only from the CO_2 compressors but also heat from the purification system chiller.

A comparison of the generation and auxiliary loads reveals that the MTR-1 steam turbine generator is nearly unchanged from the Base Case. The small change is a result of the steam extraction required for CO_2 drying. In contrast, the MEA cases are nearly 100 to 120 MW lower in gross generation because of the steam extraction for solvent regeneration. Both the MTR-1 and the MEA cases have additional generation related to the CO_2 removal system. The total gross generation for the MTR-1 case is nearly 100 MW higher than the MEA-1 case. The auxiliary load of the existing plant for the MTR-1 case is up slightly from approximately 29,700 kW to 33,430 kW. This is mostly a result of the increased fan load on the secondary air fan seeing both a 20% higher flow along with a notable increase in head. The largest change in this section is the auxiliary load for the CO_2 removal system, which for the MTR-1 case is nearly 143 MW, while the MEA-1 and -1A cases are nearly 55 MW. What the MTR case gained in increased gross generation, it lost back in an increased auxiliary load. Said differently, instead of having a large steam auxiliary like the MEA case, the MTR case has a large electric auxiliary load.

The MTR-1 Case yields a net generation of 310,542 kW, which is higher than the MEA-1 case value of 303,310 kW, but less than the Advanced MEA-1A case value of 319,280 kW. The advanced MTR membrane case discussed in Section 5.4.2 yields a net generation of 328,946 kW.

Since the coal input only varies by a single percent between all the cases, the net plant efficiency trend simply echoes that of the net plant generation.

The MTR-1 case produced about 0.7% more CO_2 than the other cases because of the decreased boiler efficiency. All capture cases capture 90% of the CO_2 produced in the boiler.

Incremental capital and O&M costs for the MTR CO_2 membrane retrofit are compared to the base case (status quo), and to the MEA-1 (SOA 2006) and MEA-1A (Advanced MEA Technology) cases in Exhibit 6-2.

Exhibit 6-2
Incremental Capital and O&M Cost Comparison for Retrofitting Conesville Unit 5

Parameter	Units	Case-0	MTR-1	MEA-1	MEA-1
			Dec 2009 $	Jul 2006 $	Dec 2009 $
Capital Costs					
Bare Erected Cost	$1,000	Base	450,588	217,189	258,382
Eng, CM, HO, Fees, etc.	$1,000	Base	42,496	87,789	103,772
Project Contingency	$1,000	Base	8,400	56,022	66,429
Process Contingency	$1,000	Base	93,000	39,094	46,357
Total Investment Cost	**$1,000**	**Base**	**594,484**	**400,094**	**474,940**
Total Investment Cost	$/kW,net	Base	1,914	1,319	1,566
Operating & Maintenance Costs					
Fixed O&M Costs	$1000/yr	Base	4,681	2,494	2,647
Variable O&M Costs	$1000/yr	Base	7,963	17,645	20,631
Levelized, Makeup Power Cost	$1000/yr	Base	65,151	62,194	68,996
CO2 byproduct Revenue	$1000/yr	Base	-	-	-
Feedstock (Natural Gas) O&M Cost	$1000/yr	Base	-	653	575

Note: Costs for MEA-1 (Y2006 USD) are based on Reference [1], p ES-3, 123, 132-134.

6.1 CHALLENGES AND DIFFICULTIES IN COMPARING RESULTS OF DIFFERENT STUDIES

Although it is of interest to compare the amine reference study [1] results to the results of the membrane process developed in this report, it is not without difficulties and potential pitfalls. Any comparison should be made carefully, and this is particularly true of analyses that are developed by different organizations and at different times.

When costs are developed within a given study, there tends to be a high degree of consistency within the cost and performance development. Said differently, the relative accuracy between the cost and performance estimates is greatest when developed by the same estimating models, and team and when developed at the same moment in time.

This section addresses several factors that should be considered if comparing the results of the aforementioned studies. Listed below are several factors which are likely to contribute to notable differences.

- Different Cost Estimating Methodology
- Different Cost Data Bases
- Different Cost Assumptions
- Varying Exchange rates for Imported Equipment
- Potential Bias of Original Design Basis.

Each of these factors will be briefly explored in the following subsection to underscore the difficulty in this type of comparison.

6.1.1 <u>Different Cost Estimating Methodology</u>

An example of how different cost estimating methodology can affect the developed cost is in the application of design margins. One analysis might base the equipment cost directly on the conceptual heat and mass balance data, while a different cost effort might apply design margins common in a more detailed design phase. This difference is methodology can be more substantial that it first appears. For example, consider the application of a 10% flow margin on turbomachinery. If the physical flow geometry and system resistance is fixed, then a 10% flow margin will correspond to a 21% margin on head, since the system pressure drop that the turbomachinery must overcome is related to the square of the mass flow. Furthermore, the power input requirement of such a piece of turbomachinery is related to the product of the flow and the head, and thus the 10% flow margin would result in a 33% power margin. If the cost of the turbomachinery can be scaled on power with an exponential factor of 0.65, this 10% flow margin (or 33% power margin) will result in a cost increase of 20%. This is just an example of how a difference methodology or how subtle unknown assumption can influence the cost.

The WorleyParsons equipment list presented in Appendix C incorporates design margins. It is not known if the amine reference study considered margins.

Furthermore, cost margins utilized by vendors when supplying budgetary quotes are typically not revealed.

6.1.2 Different Cost Data Base

Since different vendors will provide different prices, and since the same vendor will provide different prices at different times, it is obvious that different organizations will have different cost bases. Again the greatest cost estimate consistency between different options will likely come from a single source.

6.1.3 Different Cost Assumptions

Although WorleyParsons has endeavored to replicate the design and costing assumption utilized in the reference amine study, there are many that are not specified and therefore unknown. For example, in Section 5.2.5, the retrofit labor cost factor was discussed. The current study utilized a 110% retrofit cost factor on labor, while the same cost factor for the reference study is not specified. It is possible that the reference study did not apply a retrofit cost penalty. It is also possible that it applied a larger cost factor.

6.1.4 Varying Exchange Rates for Imported Equipment

In the present anlysis, WorleyParsons went out to vendors for budgetary costs of major equipment to improve the quality of the developed cost estimate. As mentioned in the technical maturity section, Section 2.6, some equipment is beyond the limits of many suppliers and the supplied equipment may come from foreign vendors. An example is the assumption that Man Turbo will supply the vacuum pump. Therefore the foreign exchange that exists at the time of the estimate is important to the total plant cost. This may not seem significant unless one considers the foreign exchange variability as presented in Exhibit 6-3 reveals a 20% variation in just 5 months. This variability in foreign exchange will complicate the comparison of costs developed at different times.

Exhibit 6-3 Daily Exchange Rates: Euros per US Dollar

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Time period shown in diagram: 1/Jan/2010 - 31/Dec/2010

6.1.5 Potential Bias of Original Design Basis

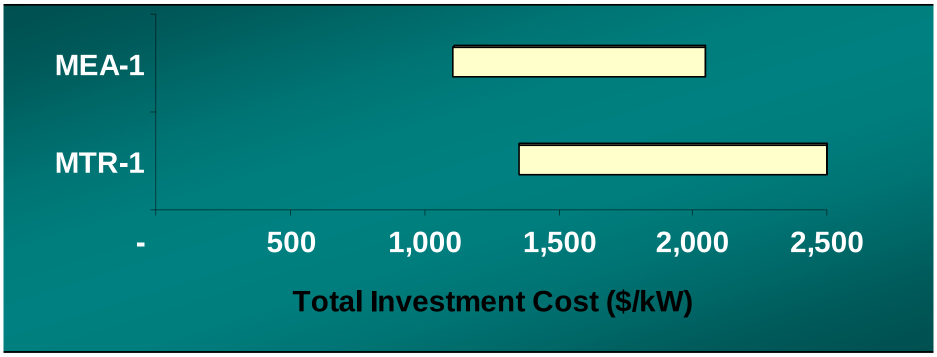
In developing the present analysis, WorleyParsons was asked to develop the analysis on a basis consistent with the amine reference study. Although this may facilitate comparison as much as possible, it is also possible to introduce an unintentional bias into the analysis. For example, the original study included an O_2 limit in the CO_2 product specification of 100 ppm. This is easy for a chemical separation process, like amine scrubbing, to achieve. However, for the membrane this subtle requirement mandates the addition of a distillation column for purification. Presently there is considerable debate over the real requirements for the CO_2 product specification. Thus, the simple replication of the original design basis may unknowingly introduce an evaluation bias.

6.1.6 Summary

In light of the above concerns, care should be taken in comparing the amine reference study to the membrane study. Likewise, care should be taken in interpreting a 20-25% cost difference as an absolute difference when each study has a cost estimate uncertainty of ±30%.

In light of the ±30% cost uncertainty, the estimate costs may be better thought of as an estimated cost range. Graphically it can be seen that the amine-based and membrane based retrofit cost estimate have overlapping costs as seen in Exhibit 6-4.

Exhibit 6-4 Total Investment Cost Overlap



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Appendix A: Heat and Mass Balance Tables

- **Boiler & Membrane**
- **Steam Turbine Cycle**

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EPRI / MTR CO2 MEMBRANE RETROFIT
Design Case

STREAM	1		2		3		4		5		6		7		8	
DESCRIPTION	Ambient Air to Primary Air Fan		Primary Air		Primary Air Leak to Secondary Air		Primary Air Leak to Flue Gas		Primary Air Air Heater Bypass		Primary Air to Coal Milling		Coal Feed		Infiltration Air	
Vapor & Liquid																
V&L Mixture Component	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %
Ar	288	0.91	288	0.91	5	0.91	61	0.91	128	0.91	222	0.91	N/A	N/A		
CO_2	9	0.03	9	0.03	0	0.03	2	0.03	4	0.03	7	0.03	N/A	N/A		
H_2O	656	2.08	656	2.08	10	2.08	139	2.08	291	2.08	507	2.08	N/A	N/A		
N_2	24,135	76.47	24,135	76.47	385	76.46	5,106	76.46	10,726	76.47	18,644	76.46	N/A	N/A		
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
O_2	6,476	20.52	6,476	20.52	103	20.52	1,370	20.52	2,878	20.52	5,002	20.52	N/A	N/A		
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
TOTAL	31,564	100.00	31,564	100.00	504	100.00	6,677	100.00	14,027	100.00	24,383	100.00	0	0.00		
V&L Mixture Component	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %
Ar	11,487	1.27	11,487	1.27	183	1.27	2,430	1.27	5,105	1.27	8,873	1.27	N/A	N/A		
CO_2	417	0.05	417	0.05	7	0.05	88	0.05	185	0.05	322	0.05	N/A	N/A		
H_2O	11,816	1.30	11,816	1.30	189	1.30	2,500	1.30	5,251	1.30	9,128	1.30	N/A	N/A		
N_2	676,111	74.54	676,111	74.54	10,798	74.54	143,024	74.54	300,464	74.54	522,289	74.54	N/A	N/A		
NO	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
NO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
O_2	207,212	22.84	207,212	22.84	3,309	22.84	43,833	22.84	92,085	22.84	160,069	22.84	N/A	N/A		
SO_2	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	N/A	N/A		
TOTAL	907,043	100.00	907,043	100.00	14,487	100.00	191,875	100.00	403,090	100.00	700,681	100.00	0	0.00		
Solid																
Solid Components	lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr	
Coal	0		0		0		0		0		0		378,425			
Ash	0		0		0		0		0		0		0			
TOTAL	0		0		0		0		0		0		378,425			
All Phases																
All Phases																
Temperature, °F	80		92		92		92		92		339		80			
Pressure, psia	14.7		15.6		15.6		15.6		15.6		15.6		14.7			
Total Flow, lb/hr	907,043		907,043		14,487		191,875		403,090		700,681		378,425			
Flow Rate, MMscfd (V only)	287		287		5		61		128		222		N/A			
Vapor Frac	1.00		1.00		1.00		1.00		1.00		1.00		N/A			
Density, lb/cuft (V & L only)	0.073		0.076		0.076		0.076		0.076		0.052		N/A			
Average MW (V & L only)	28.74		28.74		28.74		28.74		28.74		28.74		N/A			

Notes:
1. Results based on MTR Membrane Performance Received on 6/25/2010.

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EPRI / MTR CO2 MEMBRANE RETROFIT
Design Case

STREAM	9		10		11		12		13		14		15		16	
DESCRIPTION	Ambient Air to Booster Fan		Air to Countercurrent Module		Vitiated Air to Secondary Air Fan		Secondary Air to Air Heater		Secondary Air to Boiler		Flue Gas to Air Heater		Flue Gas to ESP		Flue Gas to ID Fans	
Vapor & Liquid																
V&L Mixture Component	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %
Ar	955	0.91	955	0.91							1,247	0.80	1,308	0.80	1,308	0.80
CO_2	31	0.03	31	0.03						9.06	30,532	19.47	30,534	18.68	30,534	18.68
H_2O	2,178	2.08	2,178	2.08							12,659	8.07	12,798	7.83	12,798	7.83
N_2	80,129	76.46	80,129	76.46							108,364	69.11	113,469	69.41	113,469	69.41
NO	0	0.00	0	0.00							50	0.03	50	0.03	50	0.03
NO_2	0	0.00	0	0.00							2	0.00	2	0.00	2	0.00
O_2	21,499	20.52	21,499	20.52						17.37	3,630	2.31	5,000	3.06	5,000	3.06
SO_2	0	0.00	0	0.00							323	0.21	323	0.20	323	0.20
C_3H_8	0	0.00	0	0.00							0	0.00	0	0.00	0	0.00
TOTAL	104,792	100.00	104,792	100.00	117,561	100.00	117,561	100.00	118,065	100.00	156,807	100.00	163,484	100.00	163,484	100.00
V&L Mixture Component	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %
Ar	38,137	1.27	38,137	1.27							49,810	1.04	52,240	1.05	52,240	1.05
CO_2	1,384	0.05	1,384	0.05							1,343,730	28.02	1,343,820	26.94	1,343,820	26.94
H_2O	39,230	1.30	39,230	1.30							228,053	4.76	230,552	4.62	230,552	4.62
N_2	2,244,700	74.54	2,244,700	74.54							3,035,650	63.30	3,178,670	63.73	3,178,670	63.73
NO	0	0.00	0	0.00							1,503	0.03	1,503	0.03	1,503	0.03
NO_2	0	0.00	0	0.00							92	0.00	92	0.00	92	0.00
O_2	687,948	22.84	687,948	22.84							116,153	2.42	159,986	3.21	159,986	3.21
SO_2	0	0.00	0	0.00							20,709	0.43	20,709	0.42	20,709	0.42
C_3H_8	0	0.00	0	0.00							0	0.00	0	0.00	0	0.00
TOTAL	3,011,398	100.00	3,011,398	100.00	3,539,302	100.00	3,539,302	100.00	3,553,779	100.00	4,795,700	100.00	4,987,572	100.00	4,987,572	100.00
Solid																
Solid Components	lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr	
Coal	0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		35,239		35,239		123	
TOTAL	0		0		0		0		0		35,239		35,239		123	
All Phases																
All Phases																
Temperature, °F	80		97		90		97		594		708		322		322	
Pressure, psia	14.7		16.1		14.7		15.3		14.9		14.4		14.0		13.9	
Total Flow, lb/hr	3,011,398		3,011,398		3,539,302		3,539,302		3,553,779		4,830,939		5,022,811		4,987,695	
Flow Rate, MMscfd (V only)	954		954		1,071		1,071		1,075		1,428		1,489		1,489	
Vapor Frac	1.00		1.00		1.00		1.00		1.00		1.00		1.00		1.00	
Density, lb/cuft (V & L only)	0.073		0.078		0.075		0.077		0.040		0.035		0.051		0.050	
Average MW (V & L only)	28.74		28.74		30.11		30.11		30.10		30.58		30.51		30.51	

Notes:
1. Results based on MTR Membrane Performance Received on 6/25/2010.

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EPRI / MTR CO2 MEMBRANE RETROFIT
Design Case

STREAM	17		18		19		20		21		22		23		24	
DESCRIPTION	Flue Gas to Cooler		Flue Gas to FGD		Desulpherized Flue Gas		Flue Gas from Compressor		Flue Gas to Direct Contact Cooler		Flue Gas to Cross Flow Module		Direct Contact Cooler Blowdown		Flue Gas to Countercurrent Module	
Vapor & Liquid																
V&L Mixture Component	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %
Ar	1,308	0.80	1,308	0.80	1,308	0.75	1,308	0.75					0	0.00		
CO_2	30,534	18.68	30,534	18.68	30,491	17.58	30,491	17.58					3	0.02		
H_2O	12,798	7.83	12,798	7.83	23,085	13.31	23,085	13.31					18,449	99.97		
N_2	113,469	69.41	113,469	69.41	113,469	65.43	113,469	65.43					0	0.00		
NO	50	0.03	50	0.03	50	0.03	50	0.03					0	0.00		
NO_2	2	0.00	2	0.00	2	0.00	2	0.00					2	0.01		
O_2	5,000	3.06	5,000	3.06	4,992	2.88	4,992	2.88					0	0.00		
SO_2	323	0.20	323	0.20	18	0.01	18	0.01					0	0.00		
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00					0	0.00		
TOTAL	163,484	100.00	163,484	100.00	173,415	100.00	173,415	100.00	180,091	100.00	161,636	100.00	18,455	100.00	133,650	100.00
V&L Mixture Component	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %
Ar	52,240	1.05	52,240	1.05	52,240	1.01	52,240	1.01					0	0.00		
CO_2	1,343,820	26.94	1,343,820	26.94	1,341,910	26.05	1,341,910	26.05					151	0.05		
H_2O	230,552	4.62	230,552	4.62	415,887	8.07	415,887	8.07					332,367	99.92		
N_2	3,178,670	63.73	3,178,670	63.73	3,178,670	61.71	3,178,670	61.71					8	0.00		
NO	1,503	0.03	1,503	0.03	1,503	0.03	1,503	0.03					0	0.00		
NO_2	92	0.00	92	0.00	92	0.00	92	0.00					92	0.03		
O_2	159,986	3.21	159,986	3.21	159,742	3.10	159,742	3.10					1	0.00		
SO_2	20,709	0.42	20,709	0.42	1,152	0.02	1,152	0.02					5	0.00		
C_3H_8	0	0.00	0	0.00	0	0.00	0	0.00					0	0.00		
TOTAL	4,987,572	100.00	4,987,572	100.00	5,151,195	100.00	5,151,195	100.00	5,368,289	100.00	5,035,663	100.00	332,624	100.00	3,971,805	100.00
Solid																
Solid Components	lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr	
Coal	0		0		0		0		0		0		0		0	
Ash	123		123		31		31		31		15		15		8	
TOTAL	123		123		31		31		31		15		15		8	
All Phases																
All Phases																
Temperature, °F	351		259		125		261		253		95		95		85	
Pressure, psia	15.1		15.0		14.7		29.0		29.0		29.0		29.0		27.6	
Total Flow, lb/hr	4,987,695		4,987,695		5,151,226		5,151,226		5,368,320		5,035,679		332,639		3,971,812	
Flow Rate, MMscfd (V only)	1,489		1,489		1,579		1,579		1,640		1,472		N/A		1,217	
Vapor Frac	1.00		1.00		1.00		1.00		1.00		1.00		0.00		1.00	
Density, lb/cuft (V & L only)	0.053		0.059		0.070		0.112		0.113		0.152		62.053		0.140	
Average MW (V & L only)	30.51		30.51		29.70		29.70		29.81		31.15		18.02		29.72	

Notes:
1. Results based on MTR Membrane Performance Received on 6/25/2010.

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EPRI / MTR CO2 MEMBRANE RETROFIT
Design Case

STREAM	25		26		27		28		29		30		31		32	
DESCRIPTION	Flue Gas to Heater		Flue Gas to Turbo Expander		Flue Gas to Stack		Cross Flow Permeate		Condensed Water		Raw CO_2 to First Cooler		Raw CO_2 to First Knock Out Drum		Condensed Water	
Vapor & Liquid																
V&L Mixture Component	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**
Ar	1,308	1.08	1,308	1.08	1,308	1.08			0	0.00					0	0.00
CO_2	2,045	1.69	2,045	1.69	2,045	1.69			0	0.00					0	0.04
H_2O	1,499	1.24	1,499	1.24	1,499	1.24			1,735	100.00					432	99.96
N_2	109,915	90.93	109,915	90.93	109,915	90.93			0	0.00					0	0.00
NO	33	0.03	33	0.03	33	0.03			0	0.00					0	0.00
NO_2	0	0.00	0	0.00	0	0.00			0	0.00					0	0.00
O_2	6,081	5.03	6,081	5.03	6,081	5.03			0	0.00					0	0.00
SO_2	0	0.00	0	0.00	0	0.00			0	0.00					0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00			0	0.00					0	0.00
TOTAL	120,880	100.00	120,880	100.00	120,880	100.00	27,987	100.00	1,735	100.00	26,252	100.00	26,252	100.00	432	100.00
V&L Mixture Component	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**
Ar	52,240	1.52	52,240	1.52	52,240	1.52			0	0.00					0	0.00
CO_2	90,007	2.61	90,007	2.61	90,007	2.61			0	0.00					7	0.10
H_2O	27,009	0.78	27,009	0.78	27,009	0.78			31,263	100.00					7,779	99.90
N_2	3,079,090	89.41	3,079,090	89.41	3,079,090	89.41			0	0.00					0	0.00
NO	975	0.03	975	0.03	975	0.03			0	0.00					0	0.00
NO_2	0	0.00	0	0.00	0	0.00			0	0.00					0	0.00
O_2	194,574	5.65	194,574	5.65	194,574	5.65			0	0.00					0	0.00
SO_2	17	0.00	17	0.00	17	0.00			0	0.00					0	0.00
C_3H_8	0	0.00	0	0.00	0	0.00			0	0.00					0	0.00
TOTAL	3,443,911	100.00	3,443,911	100.00	3,443,911	100.00	1,063,855	100.00	31,263	100.00	1,032,591	100.00	1,032,591	100.00	7,786	100.00
Solid																
Solid Components	**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**	
Coal		0		0		0		0		0		0		0		0
Ash		4		4		4		0		0		0		0		0
TOTAL		4		4		4		0		0		0		0		0
All Phases																
All Phases																
Temperature, °F		90		225		136		90		95		150		95		95
Pressure, psia		26.1		26.0		14.7		2.9		14.7		16.9		15.9		15.9
Total Flow, lb/hr		3,443,915		3,443,915		3,443,915		1,063,855		31,263		1,032,591		1,032,591		7,786
Flow Rate, MMscfd (V only)		1,101		1,101		1,101		255		N/A		239		235		N/A
Vapor Frac		1.00		1.00		1.00		1.00		0.00		1.00		0.98		0.00
Density, lb/cuft (V & L only)		0.126		0.101		0.066		0.019		62.059		0.102		0.107		62.020
Average MW (V & L only)		28.49		28.49		28.49		38.01		18.02		39.33		39.33		18.03

Notes:
1. Results based on MTR Membrane Performance Received on 6/25/2010.

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EPRI / MTR CO2 MEMBRANE RETROFIT
Design Case

STREAM	33		34		35		36		37		38		39		40	
DESCRIPTION	Raw CO_2 to Multi Stage Compressor		Condensed Water		Mixed CO_2 to Dehydration System		Mixed CO_2 to Stripping Column Reboiler		Mixed CO_2 to Stripping Column		Stripping Column Overheads		Overheads to Module C		Module C Permeate	
Vapor & Liquid																
V&L Mixture Component	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %	lbmol/hr	Mol %
Ar			0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00		
CO_2			2	0.27	27,858	80.68	27,855	81.56	27,855	81.56	10,083	61.58	10,083	61.58		
H_2O			794	99.72	355	1.03	0	0.00	0	0.00	0	0.00	0	0.00		
N_2			0	0.00	5,697	16.50	5,697	16.68	5,697	16.68	5,695	34.78	5,695	34.78		
NO			0	0.00	7	0.02	0	0.00	0	0.00	0	0.00	0	0.00		
NO_2			0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00		
O_2			0	0.00	599	1.73	599	1.75	599	1.75	597	3.64	597	3.64		
SO_2			0	0.00	13	0.04	0	0.00	0	0.00	0	0.00	0	0.00		
C_3H_8			0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00		
TOTAL	25,820	100.00	796	100.00	34,528	100.00	34,151	100.00	34,151	100.00	16,374	100.00	16,374	100.00	9,699	100.00
V&L Mixture Component	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %	lb/hr	Mass %
Ar			0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00		
CO_2			95	0.66	1,226,010	86.82	1,225,910	87.27	1,225,910	87.27	443,731	71.30	443,731	71.30		
H_2O			14,300	99.32	6,400	0.45	0	0.00	0	0.00	0	0.00	0	0.00		
N_2			0	0.00	159,603	11.30	159,603	11.36	159,603	11.36	159,532	25.63	159,532	25.63		
NO			0	0.00	196	0.01	1	0.00	1	0.00	1	0.00	1	0.00		
NO_2			0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00		
O_2			0	0.00	19,153	1.36	19,153	1.36	19,153	1.36	19,098	3.07	19,098	3.07		
SO_2			2	0.02	809	0.06	1	0.00	1	0.00	0	0.00	0	0.00		
C_3H_8			0	0.00	0	0.00	0	0.00	0	0.00	0	0.00	0	0.00		
TOTAL	1,024,804	100.00	14,397	100.00	1,412,171	100.00	1,404,667	100.00	1,404,667	100.00	622,362	100.00	622,362	100.00	405,275	100.00
Solid																
Solid Components	lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr		lb/hr	
Coal	0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0	
All Phases																
All Phases																
Temperature, °F	95		108		95		95		-22		-19		50		50	
Pressure, psia	15.9		126.0		407.0		402.0		392.0		384.0		379.0		126.0	
Total Flow, lb/hr	1,024,804		14,397		1,412,171		1,404,667		1,404,667		622,362		622,362		405,275	
Flow Rate, MMscfd (V only)	235		N/A		312		311		127		149		149		88	
Vapor Frac	1.00		0.00		0.99		1.00		0.41		1.00		1.00		1.00	
Density, lb/cuft (V & L only)	0.107		61.526		3.141		3.090		9.025		3.659		2.890		1.007	
Average MW (V & L only)	39.69		18.09		40.90		41.13		41.13		38.01		38.01		41.79	

Notes:
1. Results based on MTR Membrane Performance Received on 6/25/2010.

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EPRI / MTR CO2 MEMBRANE RETROFIT
Design Case

STREAM	41		42		43		44		45		46		47		48	
DESCRIPTION	Recycle Gas to Expander		Recycle Gas to Duty Recovery		Recycle Gas to Flue Gas		Stripper Column Bottoms		Pumped CO_2 to Duty Recovery		Pumped CO_2 Product		Propane to Condenser		Propane to Duty Recovery Exchangers	
Vapor & Liquid																
V&L Mixture Component	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**
Ar							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2							17,773	99.98	17,773	99.98	17,773	99.98	0	0.00	0	0.00
H_2O							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
N_2							3	0.01	3	0.01	3	0.01	0	0.00	0	0.00
NO							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2							2	0.01	2	0.01	2	0.01	0	0.00	0	0.00
SO_2							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
C_3H_8							0	0.00	0	0.00	0	0.00	18,595	100.00	18,595	100.00
TOTAL	6,675	100.00	6,675	100.00	6,675	100.00	17,777	100.00	17,777	100.00	17,777	100.00	18,595	100.00	18,595	100.00
V&L Mixture Component	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**
Ar							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
CO_2							782,177	99.98	782,177	99.98	782,177	99.98	0	0.00	0	0.00
H_2O							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
N_2							71	0.01	71	0.01	71	0.01	0	0.00	0	0.00
NO							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
NO_2							0	0.00	0	0.00	0	0.00	0	0.00	0	0.00
O_2							55	0.01	55	0.01	55	0.01	0	0.00	0	0.00
SO_2							1	0.00	1	0.00	1	0.00	0	0.00	0	0.00
C_3H_8							0	0.00	0	0.00	0	0.00	819,983	100.00	819,983	100.00
TOTAL	217,088	100.00	217,088	100.00	217,088	100.00	782,304	100.00	782,304	100.00	782,304	100.00	819,983	100.00	819,983	100.00
Solid																
Solid Components	**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**	
Coal	0		0		0		0		0		0		0		0	
Ash	0		0		0		0		0		0		0		0	
TOTAL	0		0		0		0		0		0		0		0	
All Phases																
All Phases																
Temperature, °F	50		-117		80		16		35		80		130		95	
Pressure, psia	379.0		30.1		29.1		389.0		2020.0		2015.0		184.0		177.0	
Total Flow, lb/hr	217,088		217,088		217,088		782,304		782,304		782,304		819,983		819,983	
Flow Rate, MMscfd (V only)	61		61		61		N/A		N/A		N/A		169		N/A	
Vapor Frac	1.00		1.00		1.00		0.00		0.00		0.00		1.00		0.00	
Density, lb/cuft (V & L only)	2.342		0.269		0.164		61.398		62.041		53.588		1.566		29.795	
Average MW (V & L only)	32.52		32.52		32.52		44.01		44.01		44.01		44.10		44.10	

Notes:
1. Results based on MTR Membrane Performance Received on 6/25/2010.

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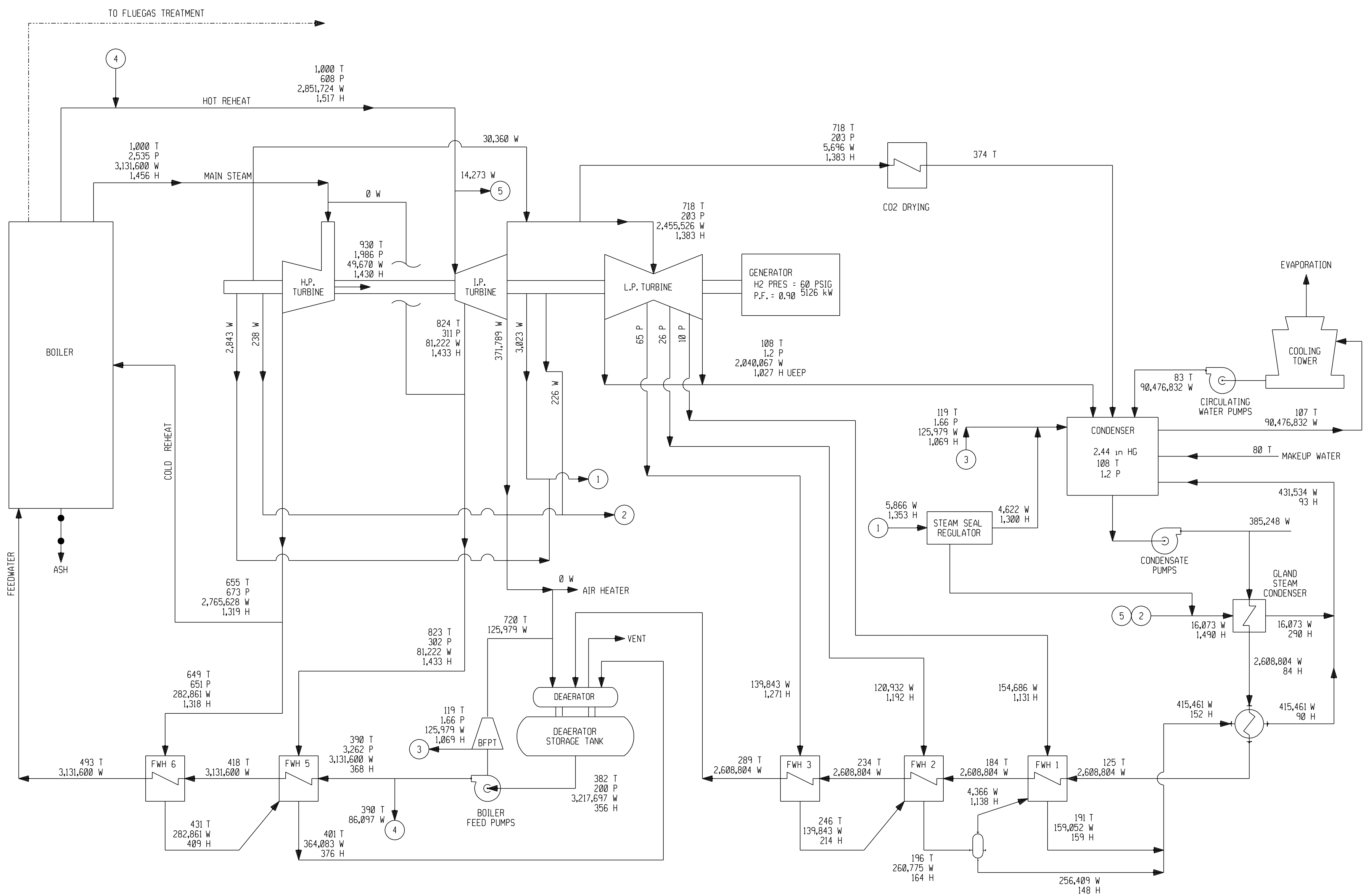
EPRI / MTR CO2 MEMBRANE RETROFIT
Design Case

STREAM	49		50		51											
DESCRIPTION	Propane to Letdown Valve		Propane to CO_2 Chiller		Propane to Compressor											
Vapor & Liquid																
V&L Mixture Component	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**	**lbmol/hr**	**Mol %**
Ar	0	0.00	0	0.00	0	0.00										
CO_2	0	0.00	0	0.00	0	0.00										
H_2O	0	0.00	0	0.00	0	0.00										
N_2	0	0.00	0	0.00	0	0.00										
NO	0	0.00	0	0.00	0	0.00										
NO_2	0	0.00	0	0.00	0	0.00										
O_2	0	0.00	0	0.00	0	0.00										
SO_2	0	0.00	0	0.00	0	0.00										
C_3H_8	18,595	100.00	18,595	100.00	18,595	100.00										
TOTAL	18,595	100.00	18,595	100.00	18,595	100.00										
V&L Mixture Component	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**	**lb/hr**	**Mass %**
Ar	0	0.00	0	0.00	0	0.00										
CO_2	0	0.00	0	0.00	0	0.00										
H_2O	0	0.00	0	0.00	0	0.00										
N_2	0	0.00	0	0.00	0	0.00										
NO	0	0.00	0	0.00	0	0.00										
NO_2	0	0.00	0	0.00	0	0.00										
O_2	0	0.00	0	0.00	0	0.00										
SO_2	0	0.00	0	0.00	0	0.00										
C_3H_8	819,983	100.00	819,983	100.00	819,983	100.00										
TOTAL	819,983	100.00	819,983	100.00	819,983	100.00										
Solid																
Solid Components	**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**		**lb/hr**	
Coal	0		0		0											
Ash	0		0		0											
TOTAL	0		0		0											
All Phases																
All Phases																
Temperature, °F	19		-31		-31											
Pressure, psia	172.0		20.0		20.0											
Total Flow, lb/hr	819,983		819,983		819,983											
Flow Rate, MMscfd (V only)	N/A		N/A		169											
Vapor Frac	0.00		0.15		1.00											
Density, lb/cuft (V & L only)	33.725		1.258		0.200											
Average MW (V & L only)	44.10		44.10		44.10											

Notes:
1. Results based on MTR Membrane Performance Received on 6/25/2010.

NOTES:



LEGEND

———	WATER OR STEAM
—●—●—	FUEL OR ASH
—·—·—	AIR
—··—··—	FLUE GAS
T	TEMPERATURE, °F
P	PRESSURE, PSIA
W	MASS FLOW, LBM/HR
H	ENTHALPY, BTU/LBM
KWe	POWER, KILOWATTS ELECTRICAL

SYSTEM PERFORMANCE SUMMARY

DRY BULB TEMPERATURE :	80 T
WET BULB TEMPERATURE :	70 T
RELATIVE HUMIDITY :	60 %
GROSS POWER:	463,044 kW

PRELIMINARY STATUS DATE — REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.

APPROVED STATUS DATE — REPRESENTS REVIEWED AND APPROVED DESIGN. ANY PORTION MARKED 'HOLD' RETAINS PRELIMINARY STATUS.

ORIGINATING PERSONNEL — PROFESSIONAL ENGINEER'S SEAL

DRAWN BY K. MCAULIFFE

CHECKED BY D. XIE

LEAD DESIGNER

ENGINEER/TECH SPECIALIST K.MCAULIFFE

PROJECT ENGINEERING MANAGER V. VAYSMAN

PROJECT MANAGER D. STAUFFER

ORIGINALLY PREPARED UNDER THE RESPONSIBLE SUPERVISION OF
PE: STATE:
LIC.NO.: DATE:

WorleyParsons
resources & energy

CLIENT/PROJECT TITLE
EPRI MTR CO2 MEMBRANE RETROFIT
AEP CONESVILLE UNIT 5
HEAT AND MASS BALANCE DIAGRAM

MTR INTEGRATION
CASE 1

SCALE NONE — DRAWING SIZE

WORLEYPARSONS DWG. NO. MTR0-DW-021-306-0001 — REV A

REV	DATE	DRAWN	CHECKED	LEAD DESIGNER	ENGINEER/TECH SPEC.	LEAD DISC ENGINEER	PROJ ENGR MANAGER	PROJECT MANAGER
A	12-8-10	KM	DX		KM			

ISSUED FOR REVIEW

FAC PIPING I & C ELECT MECH CIVIL STRUC
FAC PIPING I & C ELECT MECH CIVIL STRUC
FAC PIPING I & C ELECT MECH CIVIL STRUC
FAC PIPING I & C ELECT MECH CIVIL STRUC
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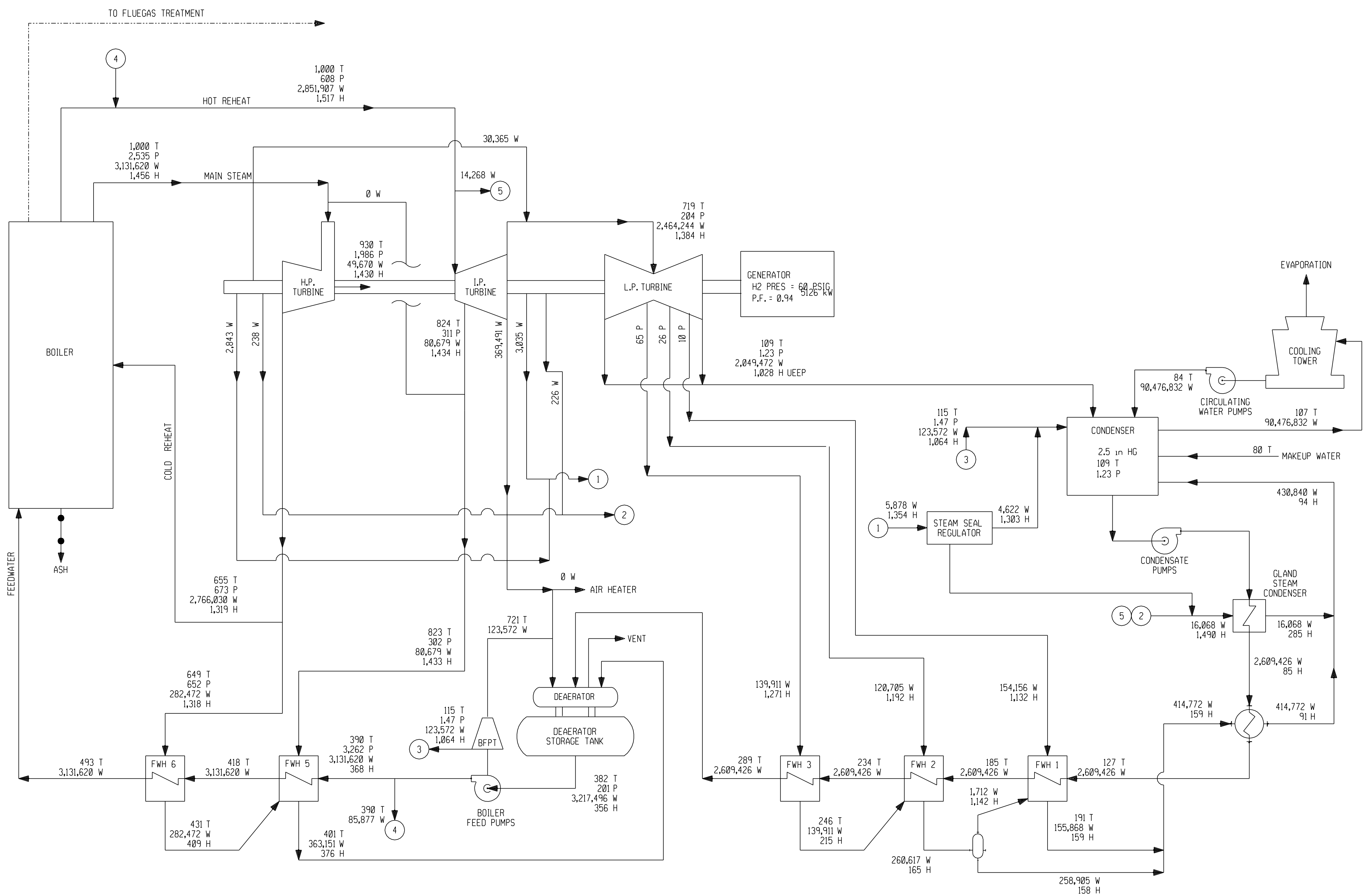
NOTES:



LEGEND

(solid line)	WATER OR STEAM
(dotted line)	FUEL OR ASH
(dash-dot line)	AIR
(dash-dot-dot line)	FLUE GAS
T	TEMPERATURE, °F
P	PRESSURE, PSIA
W	MASS FLOW, LBM/HR
H	ENTHALPY, BTU/LBM
KWe	POWER, KILOWATTS ELECTRICAL

SYSTEM PERFORMANCE SUMMARY

DRY BULB TEMPERATURE :	80 T
WET BULB TEMPERATURE :	70 T
RELATIVE HUMIDITY :	60 %
GROSS POWER:	463,478 kW

PRELIMINARY STATUS DATE — REPRESENTS GENERAL DESIGN CONCEPTS BASED ON ASSUMPTIONS. REVIEWED NOT CHECKED.

APPROVED STATUS DATE — REPRESENTS REVIEWED AND APPROVED DESIGN. ANY PORTION MARKED "HOLD" RETAINS PRELIMINARY STATUS.

ORIGINATING PERSONNEL | PROFESSIONAL ENGINEER'S SEAL

DRAWN BY K. MCAULIFFE
CHECKED BY D. XIE
LEAD DESIGNER
ENGINEER/TECH SPECIALIST K.MCAULIFFE
PROJECT ENGINEERING MANAGER V. VAYSMAN
PROJECT MANAGER D. STAUFFER

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CLIENT/PROJECT TITLE
EPRI MTR CO2 MEMBRANE RETROFIT
AEP CONESVILLE UNIT 5
HEAT AND MASS BALANCE DIAGRAM

BASE CASE
CASE 0
MAXIMUM CALCULATED - 5% OP

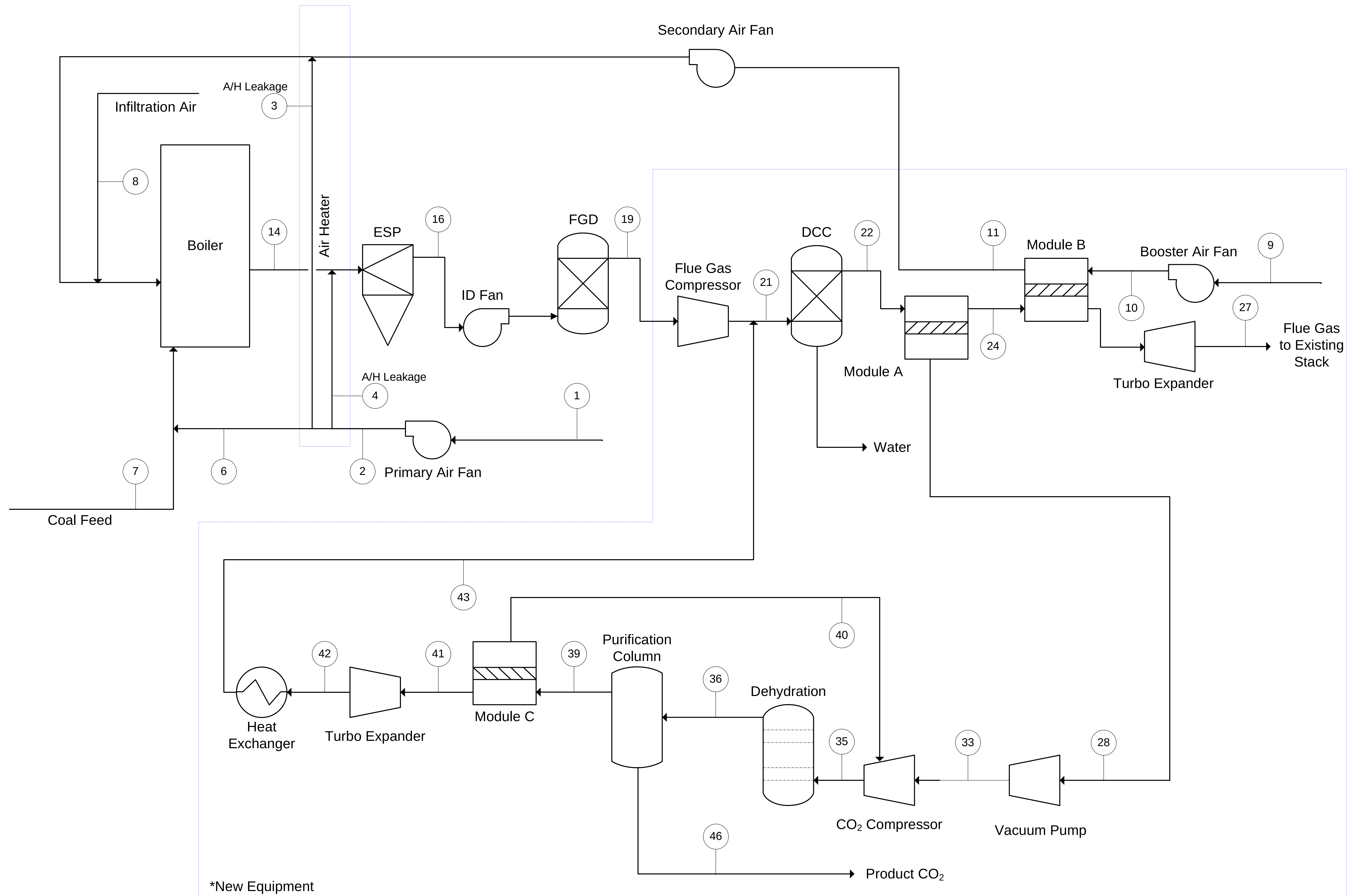
SCALE NONE | DRAWING SIZE D

WORLEYPARSONS DWG. NO. MTR0-DW-021-306-0000 | REV A

REV	DATE	DRAWN	CHECKED	LEAD DESIGNER	ENGINEER/TECH. SPEC.	LEAD DISC ENGINEER	PROJ ENGR MANAGER	PROJECT MANAGER
A ISSUED FOR REVIEW	12-8-10	KM	DX		KM			

FAC	PIPING	I & C	ELECT	MECH	CIVIL	STRUC

Appendix B: BFD & Process Flow Diagrams



NOTES:

REV	DATE	DESCRIPTION
C	12/7/2010	PROCESS UPDATE
B	11/10/2010	ISSUED FOR REVIEW
A	9/7/2010	INITIAL ISSUE

ENGINEER/TECH SPECIALIST S. O'NEILL	DATE 12/7/2010
CHECKED BY	PROJECT MANAGER D. STAUFFER

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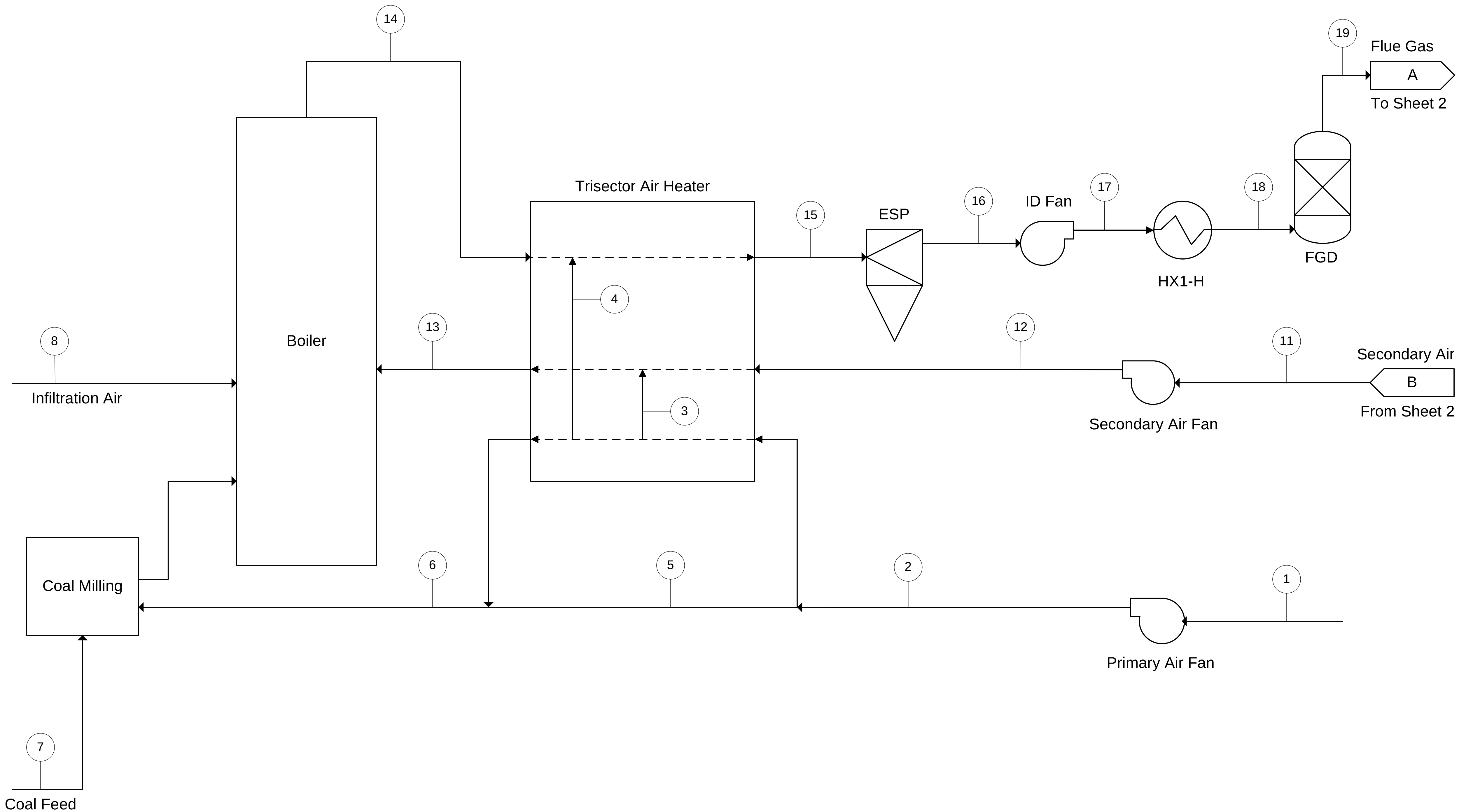
CLIENT/PROJECT TITLE

EPRI / MTR

CO2 MEMBRANE RETROFIT

BLOCK FLOW DIAGRAM

WORLEYPARSONS DWG. NO.	REV
MTR-0-DW-021-305-0101	C



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REV	DATE	DESCRIPTION
B	12/7/2010	PROCESS UPDATE
A	11/10/2010	INITIAL ISSUE

ENGINEER/TECH SPECIALIST: S. O'NEILL
DATE: 12/7/2010
CHECKED BY:
PROJECT MANAGER: D. STAUFFER

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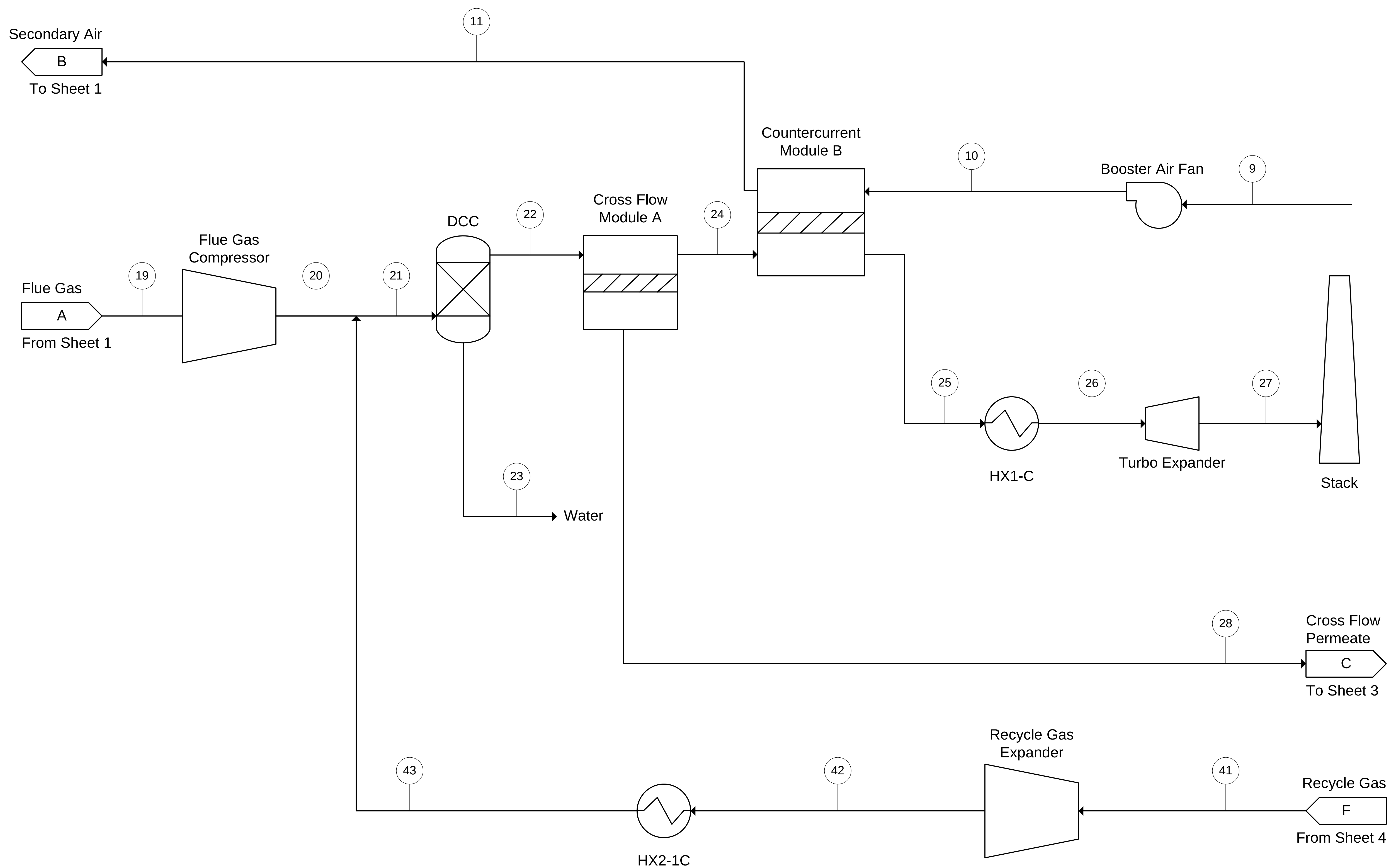
CLIENT/PROJECT TITLE
EPRI / MTR
CO2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
BOILER AIR & GAS SIDE

WORLEYPARSONS DWG. NO.: MTR-0-DW-021-305-0001
REV: B



NOTES:

REV	DATE	DESCRIPTION
B	12/7/2010	PROCESS UPDATE
A	11/10/2010	INITIAL ISSUE

ENGINEER/TECH SPECIALIST S. O'NEILL	DATE 12/7/2010
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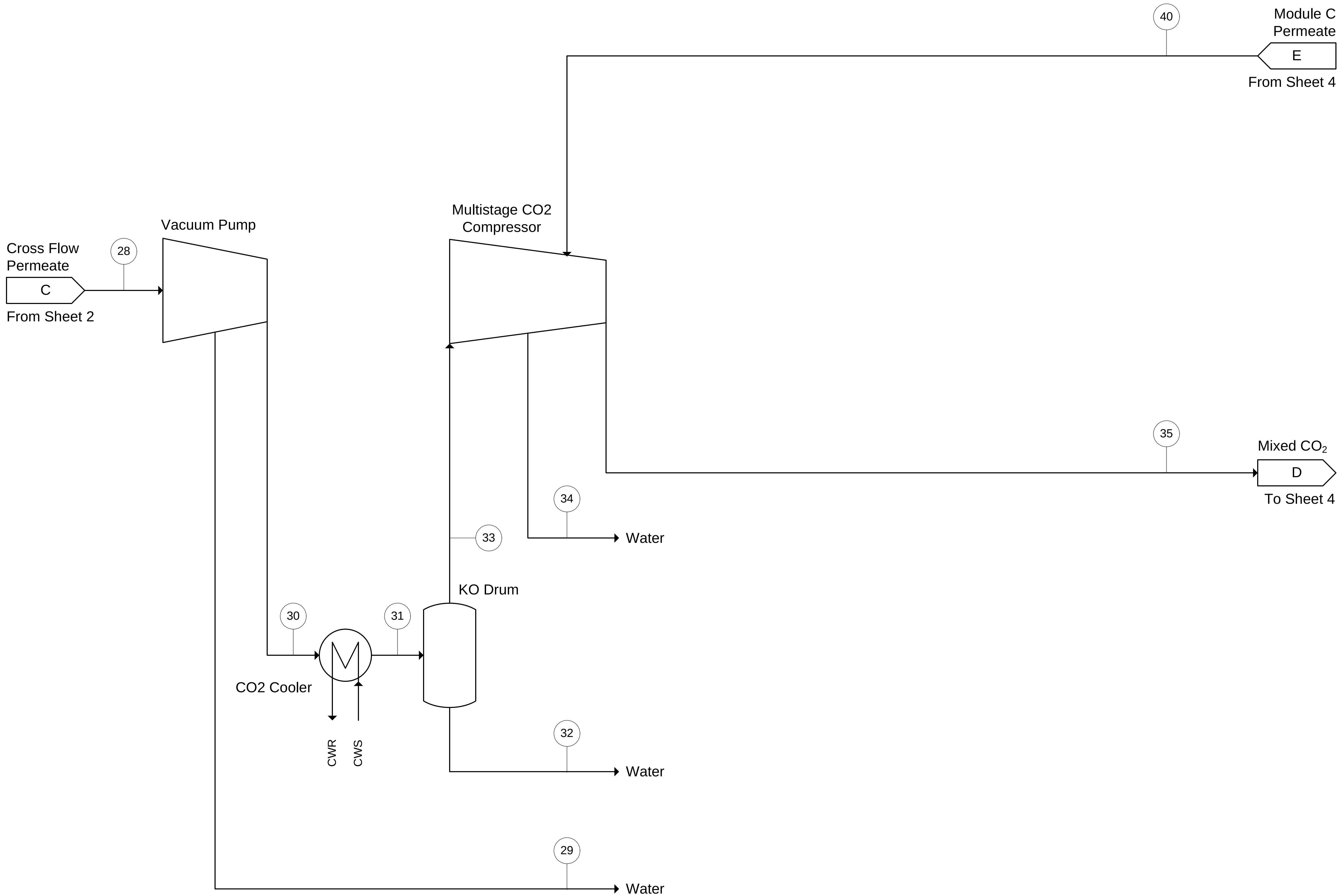
EPRI / MTR
CO2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
MEMBRANE CO2 CAPTURE

WORLEYPARSONS DWG. NO.	REV
MTR-0-DW-021-305-0002	B

NOTES:



REV	DATE	DESCRIPTION
B	12/7/2010	PROCESS UPDATE
A	11/10/2010	INITIAL ISSUE

ENGINEER/TECH SPECIALIST	DATE
S. O'NEILL	12/7/2010
CHECKED BY	PROJECT MANAGER
	D. STAUFFER

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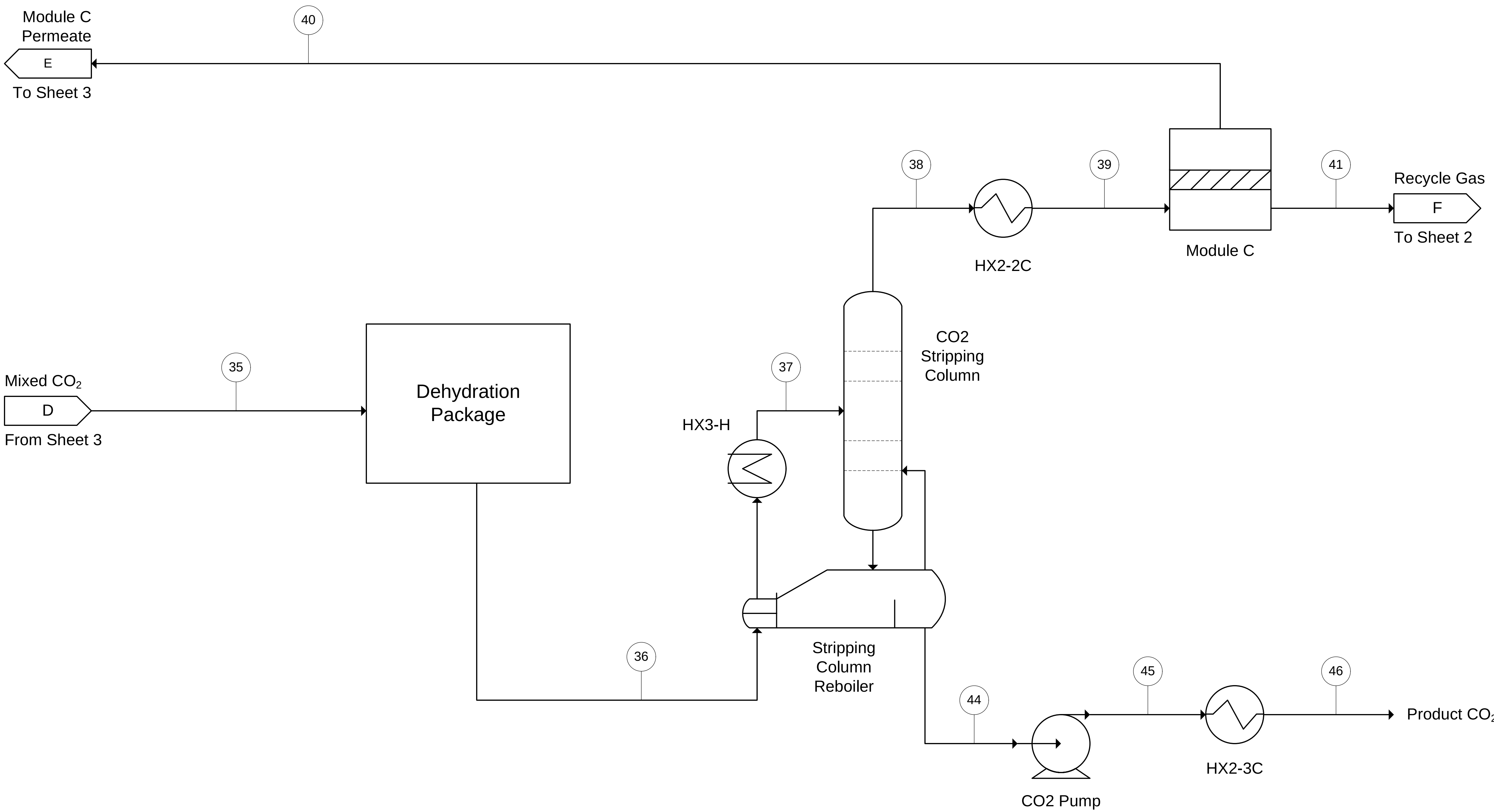
EPRI / MTR
CO2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
CO2 COMPRESSION

WORLEYPARSONS DWG. NO.	REV
MTR-0-DW-021-305-0003	B



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REV	DATE	DESCRIPTION
B	12/7/2010	PROCESS UPDATE
A	11/10/2010	INITIAL ISSUE

ENGINEER/TECH SPECIALIST: S. O'NEILL

DATE: 12/7/2010

CHECKED BY:

PROJECT MANAGER: D. STAUFFER

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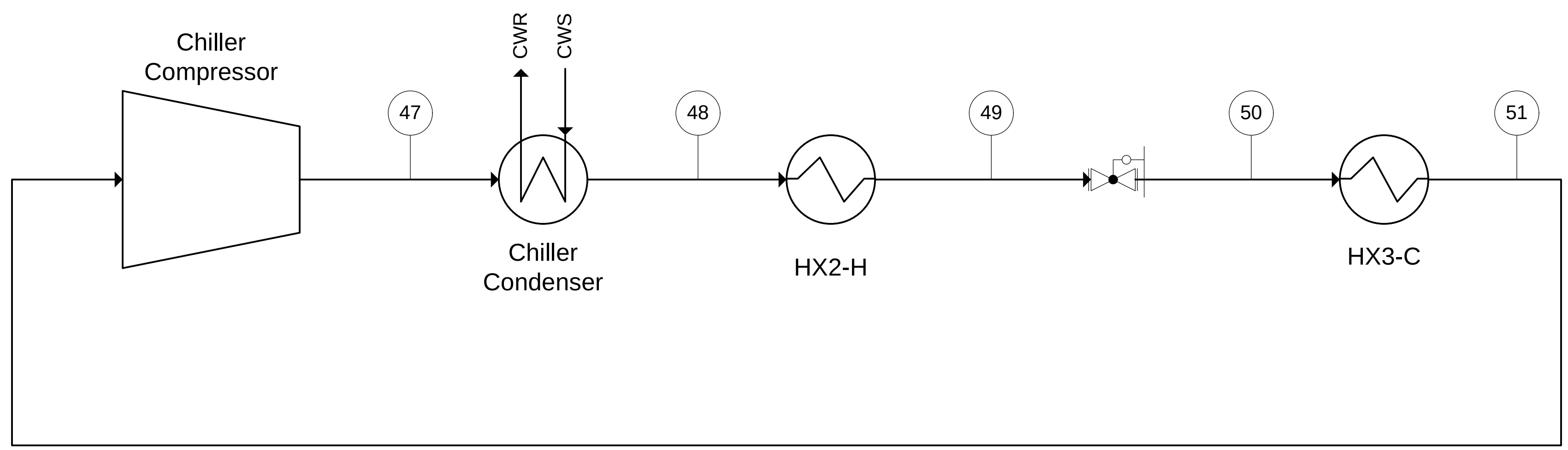
EPRI / MTR
CO2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
CO2 PURIFICATION

WORLEYPARSONS DWG. NO.: MTR-0-DW-021-305-0004

REV: B



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REV	DATE	DESCRIPTION
B	12/7/2010	PROCESS UPDATE
A	11/10/2010	INITIAL ISSUE

ENGINEER/TECH SPECIALIST: S. O'NEILL

DATE: 12/7/2010

CHECKED BY:

PROJECT MANAGER: D. STAUFFER

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CLIENT/PROJECT TITLE
EPRI / MTR
CO2 MEMBRANE RETROFIT

PROCESS FLOW DIAGRAM
CHILLER PACKAGE

WORLEYPARSONS DWG. NO.: MTR-0-DW-021-305-0005

REV: B

Appendix C: Major Equipment List

ACCOUNT 5A FLUE GAS TREATMENT EQUIPMENT

Equipment No.	Description	Type	Design Condition	Qty
1	In-situ Flue Gas / Glycol Heat Exchanger HX1-H	Finned Tube	125 MMBtu/hr 86,000 ft^2 Gas: 14 psia / 351°F Tubes: 20 psia / 295°F SS	1
2	Flue Gas Compressor	Centrifugal	339,000 acfm 14.7 psia / 29.0 psia 24,000 hp CS/SS Impeller	4
3	Direct Contact Cooler		792,000 acfm 563 MMBtu/hr Temperature: 253°F Pressure: 29.0 psia	1
4	In-situ Glycol / Flue Gas Heat Exchanger HX1-C	Finned Tube	125 MMBtu/hr 87,000 ft^2 Gas: 14 psia / 225°F Tubes: 20 psia / 295°F SS	1
5	Glycol Circulation Pump	Horizontal Centrifugal	100 ft 2,500 gpm 2 x 100%	2
6	Flue Gas Turbo Expander	Centrifugal	625,000 acfm 26 psia / 14.7 psia 42,000 hp CS/SS Impeller	1
7	Booster Air Fan	Centrifugal	757,000 acfm 14.7 psia / 16.1 psia 7,000 hp CS/CS Impeller	1

ACCOUNT 5B CARBON DIOXIDE PROCESSING

ACCOUNT 5B.1 MTR CO_2 MEMBRANE

Equipment No.	Description	Type	Design Condition	Qty
1	MTR CO_2 Module A	Cross Flow Polaris Membrane	217,000 m^2	1
2	MTR CO_2 Module B	Counter Flow Polaris Membrane	325,000 m^2	1
3	MTR CO_2 Module C	Cross Flow Polaris Membrane	4,275 m^2	1

ACCOUNT 5B.2 COMPRESSION SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	Vacuum Pump With interstage cooling	Centrifugal	522,000 acfm 2.9 psia / 16.9 psia 23,000 hp CS/SS Impeller	2
2	CO_2 Cooler	Shell and Tube	23 MMBtu/hr 50,000 ft^2 Shell: 41 psia / 109°F Tubes: 16.9 psia / 150°F CS/SS	1
3	Knock Out Drum	Vertical	14 ft dia, 18 ft T/T 15.9 psia / 95°F SS	3
4	Multistage CO_2 Compressor With interstage cooling	Centrifugal	59,000 acfm 15.9 psia / 416 psia 25,000 hp CS/SS Impeller	3
5	CO_2 Condensate Pump	Horizontal Centrifugal	100 ft 150 gpm 2 x 100%	2

Equipment No.	Description	Type	Design Condition	Qty
6	Recycle Gas Expander	Centrifugal	1,700 acfm 379 psia / 30 psia 4,000 hp SS/SS Impeller	1
7	Liquid CO_2 Pump	Horizontal Centrifugal	389 psia / 2,020 psia 861,000 lb/hr 1,700 gpm 4,000 hp 2 x 100%	2

ACCOUNT 5B.3 PURIFICATION SYSTEMS

Equipment No.	Description	Type	Design Condition	Qty
1	CO_2 Dehydration Package	TEG Dehydrator	1,550,000 lb/hr 8,200 acfm 1.0 mol % H_2O in 21 ppmv H_2O out	1
1	CO_2 Dehydration Package	Desiccant System	1,550,000 lb/hr 8,200 acfm 0.27% H_2O in -40°F dew point out	1
2	CO_2 Stripping Column Reboiler	Shell and Tube	27 MMBtu/hr 14,000 ft^2 Shell: 389 psia / 16°F Tubes: 402 psia / 95°F SS/SS	1
3	CO_2 Condenser HX3-H / HX3-C	Shell and Tube	137 MMBtu/hr 33,000 ft^2 Shell: 20 psia / -31°F Tubes: 397 psia / 32°F SS/SS	1
4	CO_2 Stripping Column	Shell and Tube	12 ft dia, 68 ft T/T 28 Trays 392 psia / -22°F SS	1

ACCOUNT 5B.4 CHILLING SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Propane Chiller Compressor	Centrifugal	75,000 acfm 20 psia / 184 psia 23,000 hp SS/SS Impeller	1
2	Chiller Condenser	Shell and Tube	138 MMBtu/hr 60,000 ft^2 Shell: 20 psia / 109°F Tubes: 184 psia / 130°F CS/CS	1
3	HX2-1C / HX2-H	Shell and Tube	11 MMBtu/hr 11,000 ft^2 Shell: 177 psia / 95°F Tubes: 30 psia / -117°F CS/SS	1
4	HX2-2C / HX2-H	Shell and Tube	12 MMBtu/hr 16,000 ft^2 Shell: 177 psia / 95°F Tubes: 384 psia / -19°F CS/SS	1
5	HX2-3C / HX2-H	Shell and Tube	20 MMBtu/hr 61,000 ft^2 Shell: 177 psia / 95°F Tubes: 2,020 psia / 35°F CS/SS	1

ACCOUNT 9 COOLING WATER SYSTEM

Equipment No.	Description	Type	Design Condition	Qty
1	Cooling Tower	Evaporative, mechanical draft, multi-cell	75°F WB 90°F CWT 115°F HWT 1,075 MMBtu/h	1
2	Circ. Water Pumps	Vertical wet pit	95 ft 47,000 gpm 2 x 50%	2
3	Cooling Tower Water Makeup Pumps	Horizontal centrifugal, double suction	100 ft 2,600 gpm 2 x 100%	2

Notes:

1. Pressure and Temperate are given at normal operating conditions.
2. Capacities, duties, and ratings are nominal design values.

Appendix D: Design Basis Document

EPRI Project Design Basis:

Design Basis Document for the MTR CO_2 Membrane for Capturing CO_2 from Power Plant Flue Gas

Draft, July 07, 2010

WorleyParsons Group Inc.
2675 Morgantown Road
Reading, PA, 19607

Project Manager
D Stauffer

NON-CONFIDENTIAL VERSION:
This document has been reviewed by WorleyParsons and MTR and does not contain any information confidential to MTR.

EPRI Project Manager
A. Bhown

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LIST OF FIGURES

LIST OF TABLES

1
SUMMARY PROCESS DESCRIPTION / OBJECTIVE

The overall objective of the WorleyParsons task is to assist EPRI in technical and economic evaluation of MTR's CO_2 membrane capture system, based on MTR's field and lab data.

The objective of the MTR CO_2 membrane capture system is the implementation of the MTR cross flow and counter flow modules with vacuum pump and sweep air respectively to accomplish CO_2 Capture from a traditional pulverized coal plant while minimizing the auxiliary load. A high level depiction of this process is presented in Figure 1-1.

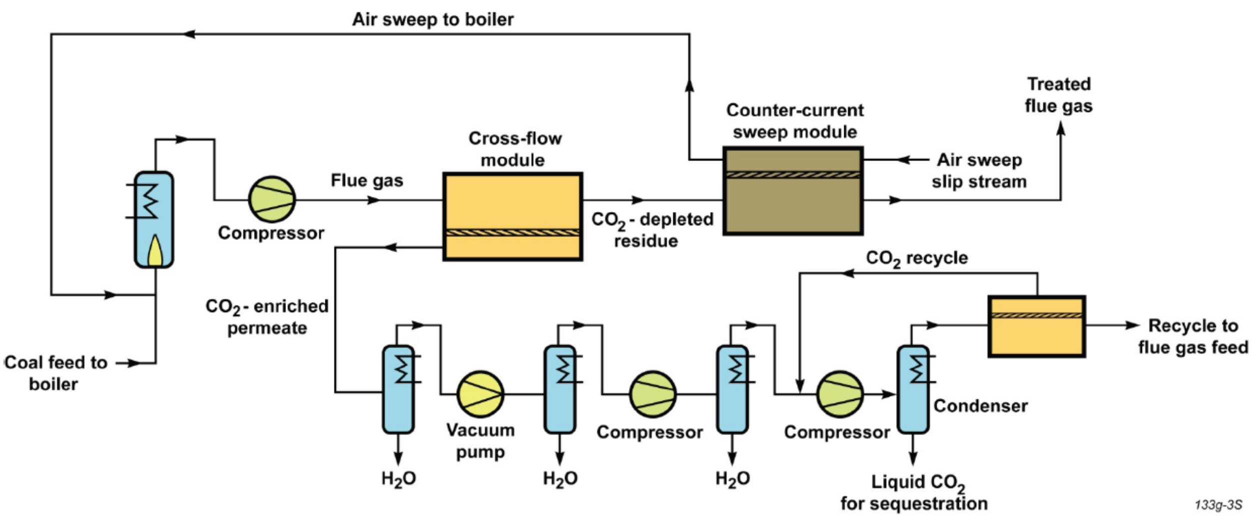


Figure 1-1
MTR CO_2 Membrane Application to a Pulverized Coal Plant – A High Level BFD

The heart of the CO_2 membrane process is the cross-flow module and the counter-current sweep module, both of which operate on a partial pressure driving force. The cross flow membrane achieves the CO_2 partial pressure driving force through a pressure gradient achieved by balancing feed compression and permeate vacuum pumps. Although vacuum pumps maybe less efficient than gas compressors, the vacuum pump contributes to the overall process efficiency since the permeate flow is a small fraction of the feed flow.

The counter-current sweep module achieves its CO_2 partial pressure driving force through the use of an air sweep stream. The advantage of using the air sweep stream is that the process air can be utilized in the PC boiler, and that the entrained CO_2 is recycled back into the process, thus increasing the overall capture rate.

The other unit operations seen in the Figure 1-1 work together to compress and purify the captured CO_2.

1.1 CO_2 Membrane Application Objectives

The objectives for MTR CO_2 membrane project include the following.

- The carbon capture basis will be 90%, based on the produced carbon dioxide. (Sec 2.7, CO_2)
- The CO_2 membrane Capture application will be based on retrofitting Conesville Unit No. 5 as documented by the DOE/NETL 401/110907 Report, entitled “Carbon Dioxide Capture from Existing Coal-Fired Power Plants, November 2007 [1, 2].

The rest of this document will present the basis of the project to be used in the process analysis.

1.2 Design Cases

A summary of the cases to be evaluated in this study is presented in.

Table 1-1
Evaluation Matrix

Case	Description	CO_2 Capture/ Compression	Cost	Notes
0	Do Nothing Case	None	NA	Existing Conesville Unit No. 5
MTR-1	MTR CO_2 Membrane Retrofit	90% capture/ 2015 psia	Dec 2009 $	Retrofit of Conesville Unit No 5 [Focus of this Evaluation.]
MEA-1	MEA Retrofit Retrofit (SOA 2006)	90% capture/ 2015 psia	Escalate to Dec 2009 $	Solvent regeneration energy of 1550 Btu/lbm-CO_2. (Note a)
MEA-1a	MEA Retrofit Retrofit (Advanced)	90% capture/ 2015 psia	Cost presumed to be equivalent to MEA-1	Solvent regeneration energy of 1200 Btu/lbm-CO_2. (Note a)

Note a. The MEA-1 and -1a Retrofit cases are known as “Case 1” and Case 1a within Reference [1].

2
DESIGN BASIS INFORMATION

2.1 Site Conditions

The conditions for the project are assumed according to the following purposes:

- Equipment design will be based on site conditions for Conesville, OH, and
- Process modeling work will be based on the American Boiler Manufacturers Association (AMBA) standard conditions [1].

The heat and mass balances will be evaluated at the AMBA standard conditions, while equipment (e.g., vacuum pumps, fans, compressors, etc.) will be sized for the indicated site conditions. This way, the performance estimate will facilitate a comparison of this study with the earlier completed Conesville Unit 5 carbon dioxide capture retrofit study [1]. The design and cost will be based on site specific conditions.

Conesville station site ambient conditions assumed in this study are based on reference [1], and presented in Table 2-1.

Table 2-1
Site Characteristics

Characteristic	Units	Value
Location		Conesville, Coshocton county, Ohio
Elevation, msl.	ft	744
Barometric Pressure	psia	14.31
Design Ambient Temperature, Wet Bulb	°F	75
Mean Coincident Dry Bulb Temperature (Note)	°F	85
Dry Bulb Temperature Extremes		
Maximum	°F	92
Minimum	°F	-1
Average cooling tower water temperature	°F	80

Note: 1% ASHRAE for Columbus, OH airport, [3]

Plant performance and heat and mass balances in this study will be referenced to the AMBA standard conditions [1] as presented in Table 2-2.

Table 2-2
AMBA Standard Conditions

Characteristic	Units	Value
Barometric Pressure	psia	14.696
Ambient Temperature, Dry Bulb	°F	80
Relative Humidity	%	60

The ambient air quality is assumed to be consistent with a dry clean air without contaminants [4], as presented in Exhibit 2-1.

Exhibit 2-1:
Ambient Air Quality

Constituent	Chemical Formula	Mole %, dry
Nitrogen	N_2	78.08%
Oxygen	O_2	20.95%
Argon	Ar	0.93%
Carbon Dioxide	CO_2	0.03%
	Total	100.00%
Impurities		
Methane	CH_4	~2 ppm
Other		Trace, (Note A)
Dust		< 0.2 mg/Nm^3

Note A: It is assumed that total content of CxHy compounds in ambient air does not exceed 9 ppm.

The following design considerations will not be quantified for this conceptual study. Allowances for normal conditions and construction will be included in the cost estimates. Typically the consideration of these factors does not have a significant impact on the cost unless the site specific situation is unusual or extreme.

- Flood plain considerations.
- Existing soil/site conditions.
- Water discharges and reuse.
- Rainfall/snowfall criteria.
- Seismic design.
- Buildings/enclosures.
- Fire protection.
- Local code height requirements.

- Noise regulations – Impact on site and surrounding area.

2.2 Coal Characteristics

An analysis of as-received mid-western bituminous coal that is currently being utilized by the Conesville Unit 5 boiler is presented in Table 2-3. [1]

Table 2-3
Design Coal

Parameter	Units	Value
Proximate Analysis		
Moisture	Wt %	10.1
Ash	Wt %	11.3
Volatile Matter	Wt %	32.7
Fixed Carbon	Wt %	45.9
Total	Wt %	100.0
Ultimate Analysis		
Moisture	Wt %	10.1
Ash	Wt %	11.3
H	Wt %	4.3
C	Wt %	63.2
S	Wt %	2.7
N	Wt %	1.3
O	Wt %	7.1
Total	Wt %	**100.0**
Higher Heating Value	Btu/lb	11,293

2.3 Sorbent

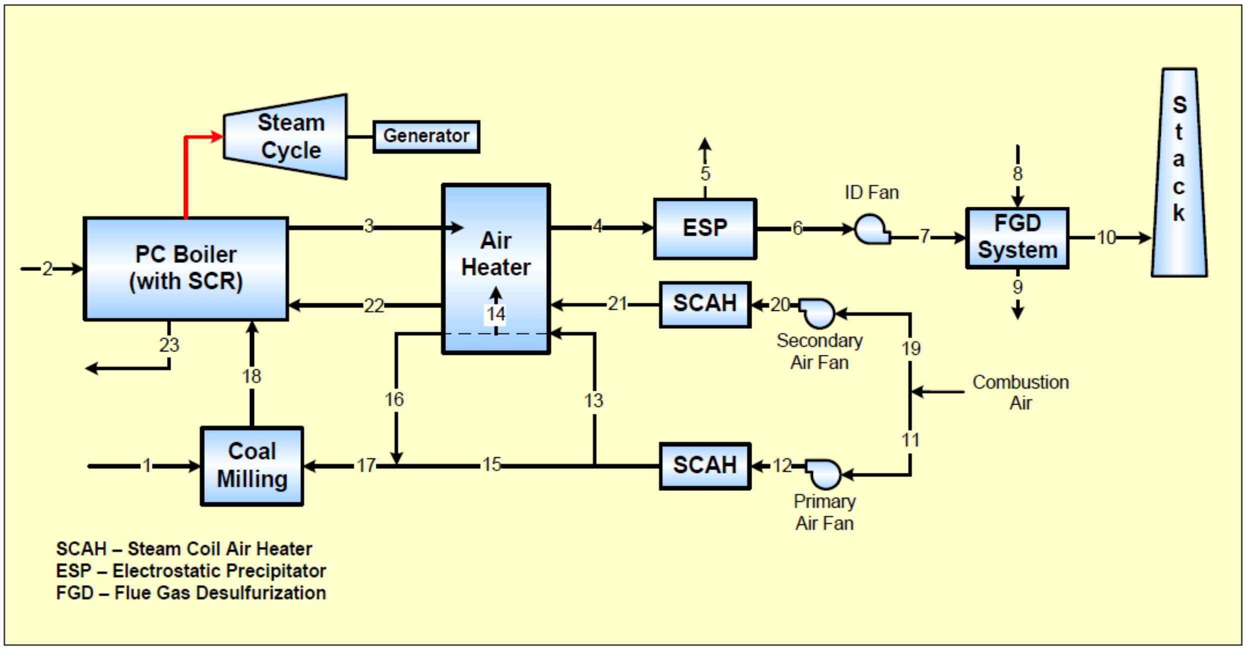
An analysis of lime that is utilized by the Conesville Unit 5 FGD system and in the reference MEA CO_2 capture study for Conesville is presented in Table 2-4. [1]

Table 2-4
Limestone Analysis

Constituent	Units	Value
CaO	wt%	90%
MgO	wt%	5%
Inerts	wt%	5%

2.4 Existing Plant Gas Side Heat & Mass Balance (Base Case)

The existing gas side configuration and heat and mass balance is shown is the block flow diagram in Figure 1-1. The corresponding material and energy balance basis is presented in Table 2-5. [1]



Material Flow Stream Identification

1	Raw Coal to Pulverizers	9	FGD System Solids to Disposal	17	Mixed Primary Air to Pulverizers
2	Air Infiltration Stream	10	Flue Gas to Stack	18	Pulverized Coal and Air to Furnace
3	Flue Gas from Economizer to Air Heater	11	Air to Primary Air Fan	19	Secondary Air to Forced Draft Fan
4	Flue Gas Leaving Air Heater to ESP	12	Primary Air to Steam Coil Air Heater	20	Secondary Air to Steam Coil Air Heater
5	Flyash Leaving ESP	13	Primary Air to Air Heater	21	Secondary Air to Air Heater
6	Flue Gas Leaving ESP to Induced Draft Fan	14	Air Heater Leakage Air Stream	22	Heated Secondary Air to Furnace
7	Flue Gas to Flue Gas Desulfurization System	15	Tempering Air to Pulverizers	23	Bottom Ash from Furnace
8	Lime Feed to FGD System	16	Hot Primary Air to Pulverizers		

Reference [1]

Figure 2-1
Simplified Gas Side Block Flow Diagram for the Existing Conesville Unit No 5 (Base Case)

Table 2-5
Gas Side Material and Energy Balance for the Existing Conesville Unit No 5 (Base Case)

Constituent	(Units)	1	2	3	4	5	6	7	8	9	10	11	12	13
O_2	(lbm/hr)	26586	42147	101097	144807		144817	144817	5335		144578	203237	203237	112918
N_2	"	4868	139626	2797385	2942220		2942220	2942220			2942220	673283	673283	374075
H_2O	"	37820	2357	228849	231294		231294	231294	250709	45979	436024	11365	11365	6314
CO_2	"			867210	867210		867210	867210			866156			
SO_2	"			20202	20202		20202	20202			1063			
H_2	"	16102												
Carbon	"	236665												
Sulfur	"	10110												
Ca	"								12452					
Mg	"								584					
MgO	"									484				
$MgSO_3$	"									1293				
$MgSO_4$	"									94				
$CaSO_3$	"									31579				
$CaSO_4$	"									2468				
$CaCO_3$	"									2398				
Ash/Inerts	"	42313		33851	33851	33851			968	968				
		Raw Coal	Leakage Air	Flue gas to AH	Flue gas to ESP	Flyash	Flue gas to ID Fan	Flue gas to FGD	Lime Slurry	FGD Disposal	Flue gas to CO_2 Sep	Pri Air to PA Fan	PA from PA Fan	Pri Air to AH
Total Gas	(lbm/hr)		184130	4014743	4205743		4205743	4205743			4390042	887885	887885	493308
Total Solids	"	374455		33851	33851	33851			14003	42884				
Total Flow	"	374455	184130	4048594	4239594	33851	4205743	4205743	270067	88863	4390042	887885	887885	493308
Temperature	(Deg F)	80	80	706	311	311	311	325	80	136	136	80	92	92
Pressure	(Psia)	14.7	14.7	14.6	14.3	14.7	14.2	15	14.7	14.7	14.7	14.7	15.6	15.6
$h_{sensible}$	(Btu/lbm)	0.000	0.000	161.831	57.924	57.750	57.924	61.384	0.000	14.116	14.116	0.000	2.899	2.899
Chemical	(10^6 Btu/hr)	4228.715												
Sensible	(10^6 Btu/hr)	0.000	0.000	655.007	245.567	1.955	243.612	258.166	0.000	3.314	63.916	0.000	2.574	1.430
Latent	(10^6 Btu/hr)	0.000	2.475	240.291	242.858	0.000	242.858	242.858	0.000	0	464.020	11.933	11.933	6.630
Total Energy[1]	(10^6 Btu/hr)	4228.715	2.475	895.298	488.425	1.955	486.470	501.024	0.000	3.314	527.936	11.933	14.507	8.060

Constituent	(Units)	14	15	16	17	18	19	20	21	22	23			
O_2	(lbm/hr)	43720	90319	66680	156999	183585	641283	641283	641283	643801				
N_2	"	144835	299208	299208	520107	524975	2124443	2124443	2124443	3122785				
H_2O	"	2445	5051	3729	8779	46599	35860	35860	35860	36001				
CO_2	"													
SO_2	"													
H_2	"					16102								
Carbon	"					236655								
Sulfur	"					10110								
Ca	"													
Mg	"													
MgO	"													
$MgSO_3$	"													
$MgSO_4$	"													
$CaSO_3$	"													
$CaSO_4$	"													
$CaCO_3$	"													
Ash/Inerts	"					42313					8463			
		Air Htr Lkg Air	Tempering Air	Hot Pri Air	Mixed Pri Air	Coal-Pri Air Mix	Sec Air to FD	Sec air to SCAH	Sec Air to AH	Hot Sec Air	Bottom Ash			
Total Gas	(lbm/hr)	191000	394577	291308	685885		2801587	2801587	2801587	2812587				
Total Solids	"										8463			
Total Flow	"	191000	394577	291308	685885	1060340	2801587	2801587	2801587	2812587	8463			
Temperature	(Deg F)	92	92	666	339		80	86.4	86.4	616.1	2000			
Pressure	(Psia)	15.6	15.6	15.6	15.6	15.0	14.7	15.2	15.1	14.9	14.7			
$h_{sensible}$	(Btu/lbm)	2.899	2.899	145.249	63.358		0.000	1.549	1.549	132.582	480.000			
Chemical	(10^6 Btu/hr)					4228.715								

Reference [1]

2.5 Existing Plant Design of Major Components

Unit 5 existing major components relevant to the CO_2 capture retrofit applications are discussed in this section and based on Reference [1]. In fact, much of the descriptive verbiage has been copied directly from this reference without a concern for copying.

Conesville Unit no 5 is owned by American Electric Power (AEP), is a coal fired steam plant that generates ~430 MWe-net using a subcritical pressure steam cycle. The plant has been in commercial operation since 1976.

This section presents details of the following existing systems/ components.

- Steam Generator (SG)
- Flue Gas Desulfurization (FGD)
- Steam Turbine Generator (STG)
- Steam Cycle Cooling
- Coal Drying

2.5.1 Existing Steam Generator

A general arrangement elevation drawing of the study unit steam generator is shown in Figure 2-2.

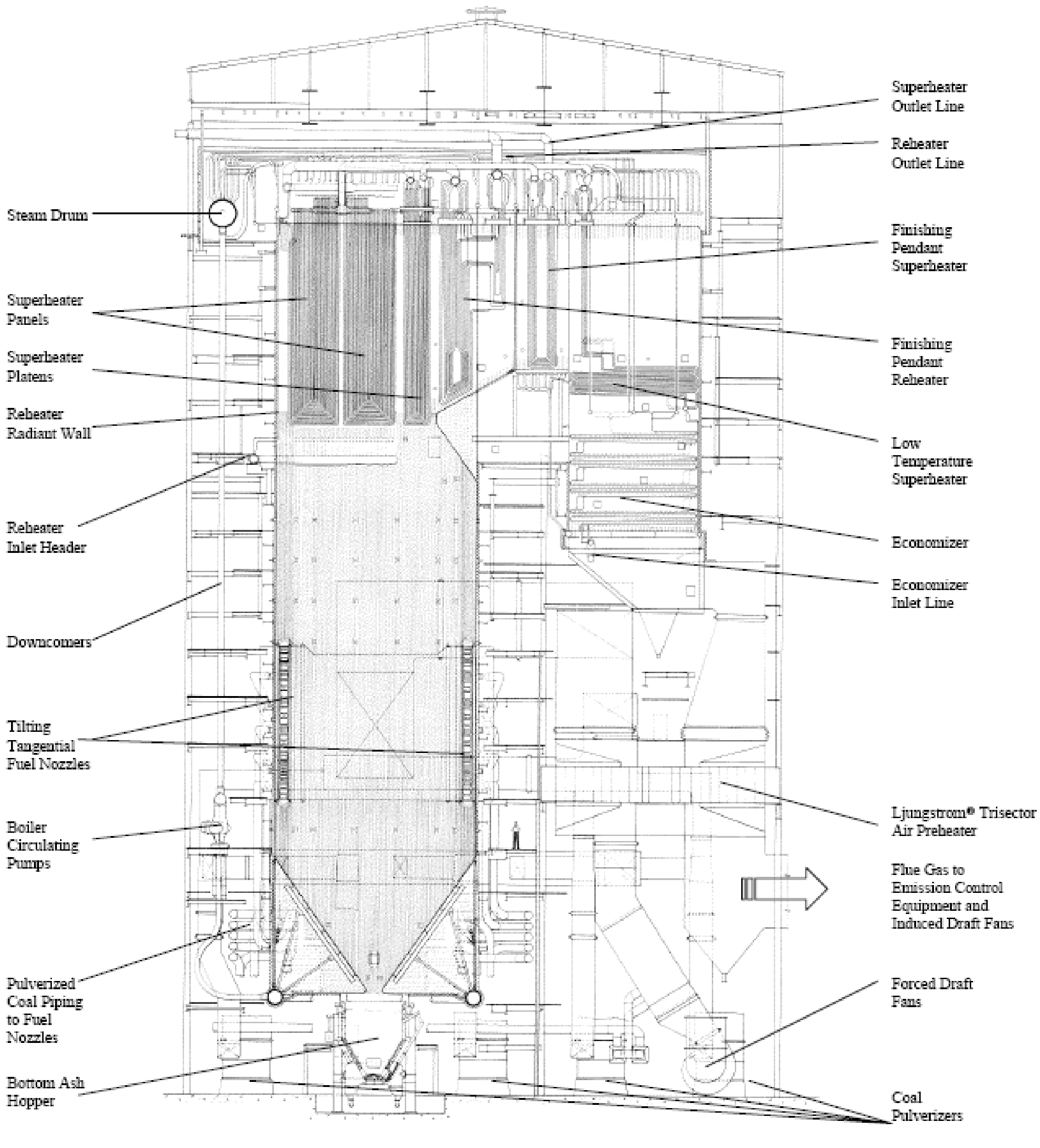


Figure 2- 1: Study Unit Boiler (Existing Conesville Unit #5 Steam Generator)

The steam generator can be described as a tangentially coal fired, subcritical pressure, controlled circulation, and radiant reheat wall unit. The furnace is a single cell design utilizing five elevations of tilting tangential coal burners. The furnace is about 15.75m (51.67 ft) wide, 13.51 m (44.33 ft) deep and 52.33 m (171.67 ft) high. The unit fires mid-western bituminous coal. The coal is supplied to the five burner elevations with five RP-903 coal pulverizers. The unit is configured in a “Conventional Arch” type design and is representative in many ways of a large number of coal-fired units in use throughout the US today. The unit is designed to generate about

391 kg/s (3.1 x 10^6 lbm/hr) of steam at nominal conditions of 175 bara (2,535 psia) and 538 °C (1,000 °F) with reheat steam also heated to 538 °C (1,000 °F). These represent the most common steam cycle operating conditions for the existing US fleet of utility scale power generation systems. Outlet steam temperature control is provided with de-superheating spray and burner tilt.

A summary of the existing Conesville Unit 5 air-fired steam generator technical information is provided in Table 2-6. [1, 5]

Table 2-6
Conesville Unit 5 Steam Generator

Existing Air Fired Boiler	**Units**	**Value**
Commercial operation	starting year	1976
Design steam output MCR	lb/hr	3,100,000
Design steam pressure	psia	2,400
Design steam temperature	°F	1,000
Design reheat temperature	°F	1,000
Furnace type		4 corner
Combustion system type		Tilting/Tangential
Burner elevations		5
Coal pulverizer model/Quantity		RP-903 / 5
Bottom ash in boiler	fractional	0.2
Residual carbon in ash	%	2.4%
Boiler outlet pressure (at air heater outlet)	psia	14.3
Primary air fan outlet pressure	in wg	TBD[a] (note a)
Primary fan efficiency, polytropic	%	TBD[a] (note a)
Secondary air fan outlet pressure	in wg	TBD[a] (note a)
Secondary fan efficiency, polytropic	%	TBD[a] (note a)
ID Fan outlet pressure	in wg	TBD[a] (note a)
ID fan efficiency, polytropic	%	TBD[a] (note a)
Economizer		
Feedwater exit temperature	°F	493
Tubes, spiral-finned	fins/inch	2
Banks quantity		4
Air Heater		
Type		Ljungstrom Tri-Sector
Flue gas exit temperature	°F	311

[a] Parameter is available through in-house confidential information. WorleyParsons is awaiting the release from AEP via DOE.

Existing Air Fired Boiler	Units	Value
ESP		
Removal efficiency	%	TBD[a] (note a)
ESP ΔP	psi	TBD[a] (note a)
Flue Gas		
Primary /secondary air split (to FD Fans)	fractional	0.241 / 0.759
Furnace pressure	in WG	-0.5
Excess air above stoichiometric	% wt	15%
Infiltration air, based on total O_2 requirement (infiltration air/ 115% air)	%	5.3%
VWO, 5% overpressure cycle		
SHO steam flow	lb/hr	3,131,619
SHO steam pressure	psia	2,535
SHO steam temperature	°F	1,005
SHO steam enthalpy	Btu/lb	1,459.7
FWI flow	lb/hr	3,131,619
FWI pressure	psia	3,165
FWI temperature	°F	496.2
FWI enthalpy	Btu/lb	483.2
ECO flow	lb/hr	3,017,507
ECO pressure	psia	3,070
ECO temperature	°F	630
ECO enthalpy	Btu/lb	652.8
RHO steam flow	lb/hr	2,850,885
RHO steam pressure	psia	590.8
RHO steam temperature	°F	1,005
RHO steam enthalpy	Btu/lb	1,517.1
RHI steam flow	lb/hr	2,765,058
RHI steam pressure	psia	656.5
RHI steam temperature	°F	607.7
RHI steam enthalpy	Btu/lb	1,290.4
RH spray	%	3.1%
RH spray flow rate	lb/hr	85,827
SH spray flow	lb/hr	114,112
SH spray flow	% of SHO	3.6%
Steam to SCAH	lb/hr	0
Burner tilts	Degree	-10
Condenser backpressure	in HgA	2.5
Boiler efficiency	%	88.13%

Additional information for the base case boiler and flue gas conditions can be found in the Reference [1] document Table 2-1.

Carbon dioxide emissions for the base case in reference [1] is 866,156 lbm/hr.

It is expected that upon retrofitting to accommodate the air sweep that the existing boiler should be able to meet required steam generation duty with the following boiler systems modifications:

- Addition of flue gas duct work associated with Air Sweep Return and heating.
- Modifications of primary and secondary air systems. [TBC]
- Other [TBD]

In order to facilitate compatibility with the reference MEA CO2 capture study performed on Conesville unit 5, it will be assumed that the unit will operate with 5% overpressure.

2.5.2 Flue Gas Desulfurization (FGD) System

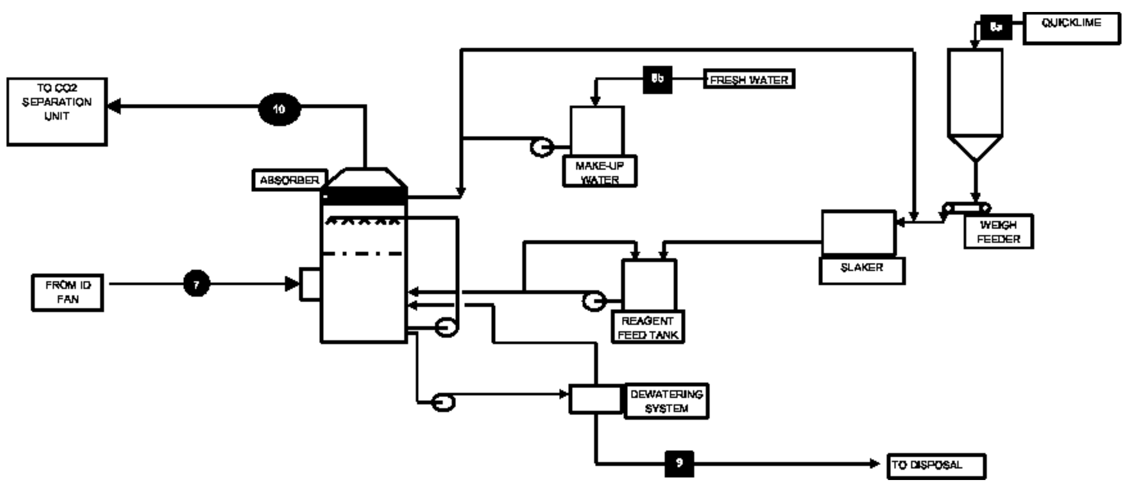


Figure 2-2
Existing Flue Gas Desulfurization System Process Flow Diagram

The existing Conesville Unit 5 is equipped with wet lime flue gas desulfurization (FGD) system to control SO2 emissions as shown in Figure 2-2. The FGD system is of natural oxidation type and generates primarily calcium sulfite waste solids for disposal. The FGD system is located downstream from an induced draft (ID) fan. A summary of the existing FGD system performance is presented in Table 2-7.

Table 2-7
Existing FGD System

Existing FGD	***Units***	***Value***
Ca/S	Mol Ratio	1.04
Solids	wt%	20%
Bypass leakage	wt%	2.50%
Liquid/Gas (L/G) Ratio	gpm/1000acfm	55
Reported SO_2 removal efficiency absorber	%	97.20%
Oxidation Oxygen/SO_2 removal ratio	mol ratio	2.3
Oxidation air pressure	psia	0.45
FGD outlet pressure	in wg	15.2
FGD outlet temperature	°F	135.0
Absorber inlet		
O_2	mol/hr	4,469
N_2	mol/hr	105,018
H_2O	mol/hr	12,853
CO_2	mol/hr	19,743
SO_2	mol/hr	315
O_2	vol%	3.14%
N_2	vol%	73.74%
H_2O	vol%	9.03%
CO_2	vol%	13.86%
SO_2	ppmv	2,212
Absorber Outlet		
O_2	mol/hr	4,461
N_2	mol/hr	105,018
H_2O	mol/hr	24,228
CO_2	mol/hr	19,720
SO_2	mol/hr	16
O_2	vol%	2.91%
N_2	vol%	68.44%
H_2O	vol%	15.79%
CO_2	vol%	12.85%
SO_2	ppmv	104

Existing FGD	*Units*	*Value*
CO_2/SO_2 Mole ratio		63
Reported absorber SO_2 removal efficiency	%	94.9%[b]

A major design criterion determining the existing FGD system SO_2 removal efficiency is the site environmental requirements. Upon conversion to the CO_2 membrane capture, controlling SO_2 for environmental purposes may not be required as most of the sulfur compounds (SO_2 and SO_3) would be co-sequestered with the CO_2.[c]

However we would expect that the FGD operation would continue at least with a water spray to provide additional particulate matter removal as well as flue gas cooling, both to precondition the flue gas prior to entering the CO_2 membrane system. Operation with a lime slurry may still be required depending on the allowable SO_2 level in the CO_2 product. Operation of the existing FGD with lime may be prudent simply to avoid having to deal with the SO_3 that would come out of the CO_2 stream during compression inter-cooling and water condensation. The lime will neutralize the H_2SO_4 that would form in either the FGD or the CO_2 compression system. It is therefore expected that the FGD operation will continue as normal, with the utilization of the lime slurry. This will also develop a sweet CO_2 product which may be required depending on the oil field being considered for EOR.

The final determination of whether the existing FGD system should continue to operate will depend on the requirements of the MTR membrane as well as the sulfur limits that may exist for the CO_2 for sequestration.

2.5.3 Steam Turbine Generator

No physical changes are expected to the steam turbine generator. Potential changes to the steam turbine cycle include the following effects of the MTR CO_2 membrane integration:

- Steam Extraction for CO_2 Dehydration process
  (If steam dehydration units are not available, natural gas fired units shall be used.)
- Heat integration of CO_2 interstage cooling with cycle condensate (parallel feedwater heating), consistent with Reference [1]

The heat and mass balance diagram for the existing Conesville Unit 5 steam turbine generator that will serve as a basis for performance if needed, is presented in Figure 2-3. [1]

[b] The absorber removal efficiency of 94.9% is known to now be better since the 2.5% weight bypass has subsequently been removed. Nevertheless, this analysis will be performed on a consistent basis with the Reference [1] study, which utilized the 2.5% bypass.

[c] SO_2 is twice as permeable as CO_2, so at 90% CO_2 capture, more than 90% of SO_2 will permeate and be captured by the membrane system. In addition, SO_3 would also permeate through the CO_2 membrane.

This turbine heat balance diagram is a valves-wide-open, 5% over pressure case utilizing a condenser pressure of 6.35 cm Hga (2.5 in.-Hga) and a steam extraction for air heating of 6.3 kg/s (50,000 lbm/hr). Per the Reference [1] study, it was assumed that this diagram reflects the design maximum allowable flow conditions of the existing turbine. Additional changes were made to this balance per below.

In order to reflect the key performance parameters of the selected unit “as designed,” heat balance diagram (Figure 2-3) was accurately re-modeled and the following adaptations to real mode operations were made:

- During normal operation no steam is required to feed the steam coil air heaters (6.3 kg/s or 50,000 lb/hr). Therefore, this extraction flow is set to zero.
- Reheat de-superheater spray water flow rate of 11 kg/s (85,827 lb/hr) is to be used as calculated in associated boiler performance computer simulation runs.

Keeping all other conditions constant, namely live steam (LS) pressure and temperature, reheat (RH) temperature and backpressure, the turbine base model reacts to the increase in RH spray (from zero to 11 kg/s or 85,827 lb/hr) and the switch-off of the extraction flow to the air reheaters (from 6.3 kg/s to 0 kg/s or from 50,000 lb/hr to 0 lb/hr) with a slight reduction in live steam flow due to the given swallowing capacity of the HP turbine (-0.26% in LS flow). In order to allow comparison with previous investigations the swallowing capacity was slightly readjusted to allow the nominal flow of 395 kg/s (3,131,619 lb/hr) at 5% overpressure.

The calculated power output applying this model showed some deficiency when compared to previous studies. This is partly due to the improved detailed modeling of the LP turbine performance, and to other differences between the previous and current models. Again, in order to allow comparison with previous investigations, the generator efficiency was adjusted in a way to allow easy comparison with previous results. Although the resulting generator efficiency may reach higher than typical values, this method allows easy comparison and simple adjustment between the two analyses, by just modifying the generator efficiency. The final steam cycle for the base case is presented in Reference [1]. The key parameters for the steam cycle reference case are listed in Table 2-8.

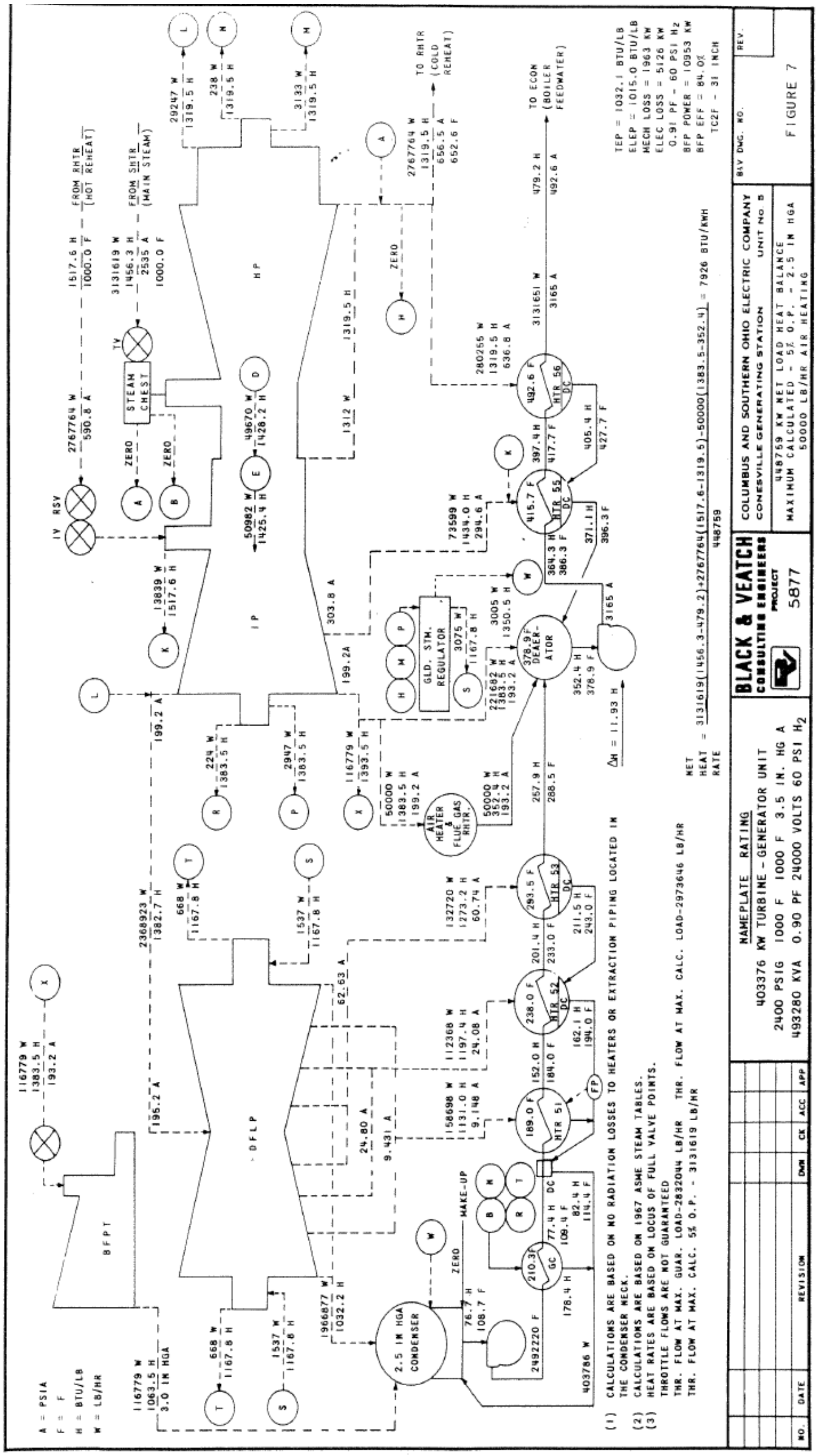


Figure 2-3
Existing Conesville Unit 5 Steam Turbine Generator Heat Balance (Basis for STG Modeling)

Table 2-8
Key Parameters of the Updated Steam Cycle (Base Case)

Live steam pressure	2,535 / 175	psia / bara
Live steam temperature	1,000 / 538	°F / °C
Live steam flow	3,131,619 / 395	lbm/hr / kg/s
Steam for air pre-heating	0 / 0	lbm/hr / kg/s
RH de-superheating spray	85,827 / 11	lbm/hr / kg/s
Backpressure	2.5 / 6.35	In. Hg abs / cm Hg abs
Power output	463,478	kW

2.5.4 Steam Cycle Cooling Strategy

The cooling tower capacity will be evaluated against the various cooling loads. The need for additional cooling capacity will be investigated during the study, (e.g., waste heat from CO_2 compression, or MTR Process related waste heat). If there is additional waste heat, it will be accommodated by either 1) the existing cooling tower, or 2) an auxiliary cooling tower if the existing cooling tower capacity is exceeded.

The existing Unit 5 utilizes a mechanical draft evaporative cooling tower and a circulating water system as a heat sink for the steam cycle and auxiliary heat loads. A summary of the existing cooling system technical information is presented in Table 2-9.

Table 2-9
Existing Cooling System

Existing Cooling System	Units	Value
Cycles of Concentration		4.0
Approach Temperature	°F	12.8
Range	°F	20
Makeup Water Flow	lb/hr	1,448
Makeup Water Temperature	°F	80

The cooling tower makeup water characteristics assumed in the study are presented in Table 2-18. This makeup water quality will allow cooling tower operation with 4 cycles[a] of concentration of dissolved solids in the circulating water.

As a part of the CO_2 membrane retrofit, a new CO_2 compression system will be added. The new CO_2 compression system will generate a substantial amount of heat that would require an additional heat sink. Utilization of the heat recovered from the CO_2 compression will reduce steam extractions from the low pressure (LP) turbine and accordingly increase steam flow rate through the last stages of the turbine. The steam turbine/condenser and circulating cooling water system's ability to accommodate increased steam flow rate are limited by their existing design. The possible limiting factors will include:

- Maximum allowable backend flow rate of the LP steam turbine. An increase of steam flow rate through the LP turbine above the design value may result in a choking condition.
- Existing condenser and circulating water systems capacities. Based on the GE design guide [6], steam turbine exhaust pressure should never exceed 5" HgA, even during transient conditions.
- Existing steam turbine generator capacity, its cooling system and electrical connections and auxiliaries.

While the possibility may exist to modify steam turbine-generator/condenser and circulating cooling systems for increased steam flow operation, such modification would require a detailed technical feasibility assessment that is not part of this study. The design cases in this study will be evaluated considering the maximum allowable flow conditions of the existing turbine, and maximum design capacities of the existing steam turbine-generator [7] and cooling water systems as presented in Table 2-10.

Table 2-10
Existing Steam Turbine-Generator and Circulating Cooling System Capacity Limits

Design Limits	Units	Value
Steam turbine LP Section maximum allowable flow (Note A)	lb/hr	1,966,877
Steam turbine IP Section maximum allowable flow	lb/hr	2,368,923
Steam turbine HP Section maximum allowable flow	lb/hr	3,131,619
Steam turbine maximum allowable exhaust pressure	in HgA	5
Generator unit maximum capacity (Note B)	kVA	493,280
Condenser maximum design duty	10^6 Btu/hr	TBD
Cooling tower maximum design duty (Note C)	10^6 Btu/hr	TBD

Notes:
A. At 2.5 in HgA exhaust pressure
B. At 0.9 PF, 24,000 V, and 60 psi H_2
C. At hottest design ambient conditions as specified in Table 2-1

Additional auxiliary cooling capacity may be required even if some of the heat from the envisioned new systems is recovered for the condensate heating. For efficient compression, CO_2 needs to be intercooled to approximately 100°F. The condensate heating system is only capable of reducing CO_2 temperature to approximately 200°F [8]. Makeup water availability and wastewater discharge limitations are amongst the major considerations when selecting a cooling system chosen, i.e. evaporative, dry, or combination of the above.

All the design cases in this study will be evaluated assuming that a sufficient amount of additional makeup water is available and that additional wastewater discharges are permitted. Hence, it is assumed that the additional auxiliary cooling system will be based on an evaporative mechanical cooling tower.

2.5.5 Coal Drying Strategy

The existing Unit 5 boiler utilizes a direct-fired pulverizing system for coal grinding, drying, pulverizing and transporting to the burners. Coal in-mill drying and transportation is accomplished by primary air preheated to ~670°F in a regenerative Ljungstrom Tri-Sector air preheater. The design bituminous coal at the Conesville site has a relatively low total moisture content of 10.1%, not requiring additional pre-drying.

During the evaluation, consideration will be given to the suitability of sweep air for use in either the primary or secondary air system. Primary air is utilized in the pulverizing system for coal drying, pulverizing and transportation functions. It is our expectation that the air sweep will be reduced in O_2 level and increased in moisture and CO_2. Under these conditions, the drying capability and temperature control for use as primary air in the existing pulverizers will have to be evaluated. Use of air sweep as the secondary is anticipated to have fewer issues, but will also need to be evaluated. [Additional discussion involving the use of sweep air is presented in Section 2.11.2.]

2.6 Environmental

The intent of the CO_2 membrane retrofit project is **use the existing Unit 5 emission control technology** rather then invest in new “BACT” technology. The incorporation of the CO_2 membrane technology is expected to eliminate the emission of 90% or more of the Carbon Dioxide and is not expected to increase any individual emission. This approach seems reasonable as the CO_2 membrane retrofit it is not expected to trigger the Prevention of Significant Deterioration (PSD) or New Source Review (NSR).

This basis appears to be consistent with the Reference [1] analysis, although it does not appear to be explicitly stated. Furthermore, the reference study only reports the CO_2 and SO_2 emissions and not emissions for carbon monoxide (CO), nitrogen dioxide (NO_2) or particulate matter (measured as PM-10 and PM-2.5), nor any other pollutants such as lead, mercury, fluorides, sulfuric acid mist, VOC, or other hazardous material.

For reference, the following background information on the Conesville plant is offered.

The Conesville Generating Station is located near Conesville, Coshocton County, in the state of Ohio. Coshocton County is located in the Zanesville-Cambridge Intrastate Air Quality Control Region (AQCR-183) and is currently designated by EPA (40 CFR 81.336) as an attainment area for all criteria pollutants (however, a portion of Coshocton County [Franklin Township] is designated non-attainment for PM-2.5); that is, the air quality in Coshocton County meets, or exceeds National Ambient Air Quality Standards (NAAQS).

Again, no environmental emission limits will be imposed on this retrofit application, other than the presumption of the continuance of the existing control equipment.

2.7 CO2 Specification

The CO_2 product specification documented in the Reference Study [1] document, which ultimately is based on the Dakota Gasification Company product specification for EOR (Dakota, 2005) is shown in Table 2-11. To facilitate comparison of the CO_2 membrane project to the amine retrofit, it is prudent to utilize this basis if possible.

Table 2-11
Dakota Gasification Project's CO2 Specification for EOR

Component	(units)	Value
CO2 Product Utilization		EOR
Pressure	Psia	2015
CO_2	(vol %)	96
H_2S	(vol %)	1
CH_4	(vol %)	0.3
C_2 + HC's	(vol %)	2
CO	(vol %)	---
N_2	(vol %)	0.6
H_2O	(ppm by vol.)	2
O_2	(ppm by vol.)	100
Mercaptans and other Sulfides	(ppm by vol)	300
SO2		NS

NOTE: This Table is from Reference [1], Table 1-1, as well as Table -3-4.

The CO_2 product is consistent with use in enhanced oil or gas recovery or for sequestration. A CO_2 product pressure of 139 bara (2,015 psia) was utilized for the amine reference study as will be used in the membrane case.

The information above is not complete for a CO_2 Product specification for combustion based sources. The Dakota Gasification Company EOR application is gasification based, thus the limit for sulfur is expressed as H_2S. For a combustion based plant, the limit would be better expressed as SO_2 or possibly as SO_2 and SO_3.

There are other possible carbon dioxide specifications that could be implemented. For maximum consistency with the reference study, we shall examine other DOE/NETL specifications. Table 2-12 presents the DOE/NETL CCS system analysis guidelines issued in 2005. [9] This specification lists neither a H_2S or SO_2 requirement, but lists a lower oxygen level (40 ppm) and a higher product pressure level (2204 psi/ 152 bars).

Table 2-12
Alternate CO2 Specification – DOE/NETL CCS System Analysis Guidelines - 2005

2.4.1 Carbon Dioxide

Carbon dioxide (CO_2), whether being sold for chemical processing or being sequestered, is to be supplied as a liquid and must meet the pipeline specification shown in Table 4 (Bock et al, 2002:

Table 4. Carbon Dioxide Pipeline Specification

Pressure	152 bar
Water Content	233 K (-40 °F) dew point
N_2	<300 ppmv
O_2	< 40 ppmv
Ar	<10 ppmv

Reference: **[9]**

Another DOE/NETL reference (CCS Systems Analysis, Technical Note 10, 2007) lists yet another set of CO_2 product specifications depending upon the location and use of the CO_2 sequestration in relation to the source. [10] This reference presents both H_2S and SO_2 levels, with SO_2 levels of <40 ppmv and <3% for EOR or Geologic sequestration respectively, and an acceptable oxygen level of either <40 or <100 ppm for EOR or Geologic sequestration respectively. The CO_2 product pressure is specified as either 2200 or 1600 psia depending upon whether the sequestration site is remote or adjacent to the source.

Table 2-13
Alternate CO_2 Specification -DOE/NETL CCS System Analysis Technical Note 10 - 2007

Table 2 - Recommended CO_2 Sequestration-Gas Design Basis

	Design Condition 1	Design Condition 2	Design Condition 3	Design Condition 4
	Remote EOR	Adjacent EOR	Remote Geological	Adjacent Gelogical
Pipeline material	carbon steel	carbon steel	carbon steel	304/316 SS
Compression pressure (psia)	2200	1600	2200	1600
CO_2	>95 vol%	>95 vol%	not limited[1]	not limited[1]
Water	dehydration[2] (0.015 vol%)	dehydration[2] (0.015 vol%)	dehydration[2] (0.015 vol%)	no dehydration[3] no free water
N_2	<4 vol%	<4 vol%	not limited[1]	not limited[1]
O_2	<40 ppmv	<40 ppmv	<100 ppmv	<100 ppmv
Ar	< 10 ppmv	< 10 ppmv	not limited	not limited
NH_3	<10 ppmv	<10 ppmv	not limited	not limited
CO	< 10 ppmv	< 10 ppmv	not limited	not limited
Hydrocarbons	<5 vol%	<5 vol%	<5 vol%	<5 vol%
H_2S	<1.3 vol%	<1.3 vol%	<1.3 vol%	<75 vol%
CH_4	<0.8 vol%	<0.8 vol%	<0.8 vol%	<4.0 vol%
H_2	uncertain	uncertain	uncertain	uncertain
SO_2	<40 ppmv	<40 ppmv	<3 vol%	<3 vol%
NOx	uncertain	uncertain	uncertain	uncertain

1: These are not limited, but their impacts on compression power and equipment cost need to be considered.
2: Dehydration process, such as a glycol absorber, is required.
3: Dehydration process is not required, but no free water must occur in the gas

Reference: [10]

Based on the dates of issue, it is presumed that the DOE/NETL Recommended CO_2 Sequestration gas basis (2007) listed in Table 2-13 supersedes that of Table 2-12 (2005). For interest, we will compare the Conditions in Table 2-11 and Table 2-13 for the EOR disposition since the Reference Study is based on EOR. This comparison is presented in Table 2-14. The CO_2 specification to be utilized in this study is presented in Table 2-15.

Table 2-14
CO_2 Specification Comparison and Comments

Parameter	DOE/NETL	Conesville MEA [i.e., Dakota Basis]	This Study	Notes
End Use	EOR	EOR	EOR	DOE/NETL Ref also has specs for Geologic.
Proximity	Remote/Adjacent	Not Specified	Not Specified	Influences Product Pressure Only
Pipeline material	Carbon steel	Not Specified	Not Specified	Outside of cost basis, except for other limits.
Product P (psia)	2200 / 1600	2015	2015	Arbitrary, yet important to match Ref Study.
CO_2	>95%	>96%	>96%	Largely immaterial. Expect >99%.
Water	150 ppmv	2 ppmv	150 ppmv	2 ppmv is average not spec. See Note *.
N_2	<4%	<0.6%	<0.6%	Largely immaterial. Expect <0.1%
O_2	<40 ppmv	<100 ppmv	<100 ppmv	Choose to match reference study.
Ar	<10 ppmv	Not Specified	Not Specified	Choose to match reference study.
NH_3	<10 ppmv	Not Specified	Not Specified.	Immaterial. Expect no NH_3 after FGD
CO	< 10 ppmv	Not Specified	Not Specified	Choose to match reference study.
Hydrocarbons	<5%	Not Specified	Not Specified	Not relevant to combustion case
H_2S	<1.3%	<1.0%	Not Specified	Not relevant to combustion case. See SO_2
CH_4	<0.8%	Not Specified	Not Specified	Not relevant to combustion case
H_2	Uncertain	Not Specified	Not Specified	Not relevant to combustion case
SO_2	<40 ppmv*	Not Specified	<1%	SO_2 is relevant for combustion case. The <1% H_2S is equivalent to <1% SO_2 on a sulfur basis.
NOx	Uncertain	Not Specified	Not Specified	Uncertain. Matches reference study.

Note: * The CO_2 specification cited in the Reference Study (Conesville Amine study) is based the CO_2 "specification" for the Dakota Gasification Company. Actually, the DGC "specification" is an average of "typical" historical data. [11] The DGC average moisture (2008) is 20 ppmv as opposed in the 2 ppmv (2005). Furthermore, as an average, the values are not bounding specifications. The Reference Study moisture level of 2 ppmv is a result of the Rectisol unit utilized by DGC. The 2 ppmv is not a requirement. The Reference Amine Study utilized liquefaction as a CO_2 purification method, which would utilize a molecular sieve. As such, the 2 ppmv moisture level was achievable. For the present study, a similar purification scheme is likely and 2 ppmv moisture would also be achievable. However, to allow freedom to entertain other options, the moisture specification will be set that that found in the DOE/NETL technical note No. 7 of 150 ppmv. For information, TEG dehydration would achieve circa 100 ppmv.

** The SO_2 level would be <3% vol for geologic sequestration.

Table 2-15
CO_2 Product Specification for the MTR Membrane Project Evaluation

Parameter	DOE/NETL	Conesville MEA [i.e., Dakota Basis]	Current Study Basis	Notes
End Use	EOR	EOR	EOR	
Product P (psia)	2200 / 1600	2015	2015	Matches Reference Study.
CO_2	>95%	>96%	>96%	Matches Ref Study. Expect >99%.
Water	150 ppmv	2 ppmv	150 ppmv	2 ppmv moisture for the Dakota plant is based on the use Rectisol. The DOE/NETL value of 150 ppmv is consistent with TEG dehydration which can achieve circa 100 ppmv. There is no reason to require the 2 ppmv level dehydration.
N2	<4%	<0.6%	<0.6%	Largely immaterial. Expect <0.1%
O2	<40 ppmv	<100 ppmv	<100 ppmv	Choose to match reference study.
SO2	<40 ppmv*	Not Specified	<1%	SO_2 is relevant for combustion case. The <1% H_2S is equivalent to <1% SO_2 on a sulfur basis.

Note: * The SO2 level would be <3% vol for geologic sequestration.

The carbon reduction efficiency is a nominal 90 percent removal based on carbon input from the coal and excluding residual carbon that exits the boiler with the ash. An alternate way of describing this CO_2 capturing basis, is simply 90 percent removal of the produced carbon dioxide. This basis is consistent with the calculations implied by the Reference [1] and was confirmed during the kickoff call [2].

2.8 Water

Various design conditions and characteristics are presented for cooling water and raw water per the tables below. Cooling Water is presented in Table 2-16.

Table 2-16
Cooling Water Conditions (Existing Cooling Towers)

CW Supply	Pressure at B.L. (Psia)	Temperature °F
Minimum	60	70
Normal	65	80
Maximum	90	95
Mechanical Design	150	150

CW Return	Pressure at B.L. (Psia)	Temperature °F
Minimum		100
Normal	45	110
Maximum		135
Mechanical Design	150	175

Condensate from the surface condenser is available at conditions presented in Table 2-17.

Table 2-17
Condensate (for amine make-up & possibly heat recovery)

Property	Pressure at B.L. (Psia)	Temperature °F
Normal	135	110
Mechanical Design	175	200

Fresh water is distributed for general use at hose stations. The source of this water is the clarifier, which is used for cooling tower make-up. The capacity of the existing clarifier is sufficient for make up. Its quality is as presented in Table 2-18:

Table 2-18
Raw Water (fresh water)

Components	Unit	Specifications
Si	ppm	22
Iron (as Fe)	ppm	0.18
Copper (as Cu)	ppm	0.05
Suspended Solids	ppm	15
Chlorine	ppm	100-180
Alkalinity	ppm	100
Na	ppm	100

Potable water comes from public network for safety showers and eye washes and requirements are defined by Table 2-19.

Table 2-19
Potable Water

Property	Pressure at B.L. (Psia)	Temperature °F
Normal	115	Ambient
Mechanical Design	150	150

2.9 Capacity Factor

Per Reference [1], the expected annual operating time is 7,446 hr/yr and is consistent with an 85% capacity factor.

2.10 Basis of MEA Case

The base amine CO_2 scrubbing case against which the MTR CO_2 membrane option will be compared is the Reference [1] Case 1 Option that is based on 90% CO_2 capture with a solvent regeneration energy of 1550 Btu/lb CO_2. [12] This is the state-of-the-art (SOA) advanced Amine case designed in 2006, which is 34% more efficient that the state-of-the-art amine case from 2000. [1] WorleyParsons will escalate the cost information in the reference document for Case 1. EPRI will utilize the original performance information and the escalated cost data for a performance and economic comparison to the MTR CO_2 membrane option.

It has been suggested that the Reference Study Case 1a (90% CO_2 capture with a regeneration energy of 1200 Btu/lb CO_2) could be used as a future State-of-the-Art amine cycle benchmark for the MTR CO_2 membrane case. The comparison to this Case 1a Amine case is worthy of the following notes:

- Case 1a is a sensitivity case and is not one of the fully costed cases in the reference study. Since the performance is simply presumed and not based on known technological advances, the capital costs were not developed in the study. In the reference study, Case 1a is based on the presumed performance of 1200 Btu/lb CO_2 regeneration energy and an identical capital cost as Case 1. (Ref Study [1], p150, 5th bullet) This is clearly an aggressive comparison, since the performance improvements will cost more to implement.

- Possible process improvements that could contribute to the envisioned performance include an increased use of heat exchangers, increased solvent concentration, added inhibitors, advanced amines, mechanical vapor recompression (MVR) or other performance improving features. All improvements may increase auxiliary electric power requirement and would likely increase the capital cost.

- Since the cost for Case 1a is not developed in the reference study, and since the Case 1 cost presumed applicable to Case 1a is clearly low for Case 1a, care should be used in the comparison of Case 1a to the MTR CO_2 membrane.

- It has also been suggested that the advanced SOA amine case utilize a regeneration energy of something between 1200 to 1300 Btu//lb CO_2. The selection of the precise regeneration number is not relevant to WorleyParsons scope and is deferred to MTR and EPRI.

Since WorleyParsons is tasked with escalating the MEA cost in the reference study, and since the Case 1a cost is presumed to be equivalent to that of Case 1, WorleyParsons will simply escalate the Case 1 cost. EPRI will perform the appropriate sensitivity study and document it with the appropriate caveats.

2.11 Technical Maturity

1. This study will be based on technology that is presently technically feasible, but not necessarily available as commercially offered equipment. Bringing the required equipment to the commercial market could require some development by an OEM. DOE/NETL may or may not require these costs to be reflected in the cost estimate. DOE/NETL may not require Non-Recurring Engineering (NRE) costs to be included in the cost estimate of this study.

2.12 Balance of Plant Assumptions

Assumed balance of plant requirements are presented in Table 2-20.

Table 2-20
Balance of Plant Assumptions

Plant Distribution Voltage	
Motors below 1 hp	110/220 volt
Motors 250 hp and below	480 volt
Motors above 250 hp	4,160 volt
Motors above 5,000 hp	13,800 volt
Steam Turbine generators	24,000 volt
Water and Waste Water	
Makeup Water	The water supply is assumed to be from groundwater, and is assumed to be in sufficient quantities and permitted to meet increases in plant makeup requirements. Makeup for potable water will be drawn from municipal sources
Storm Wastewater	Storm water that contacts equipment surfaces is collected and treated for discharge. Increased storm water treatment duty should require modifications to the existing facility, but will be acceptable by the existing discharge permit.
Water Discharge	Most of the wastewater is recycled for plant needs. Blowdown will be treated for chloride and metals, and discharged. Potential increase in blowdown should require modifications to the existing treatment system, but will be acceptable by the existing permit

3
COST BASIS INFORMATION

The basis for capital and operating costs estimates in this study will be consistent with the basis in the 2007 DOE/NETL Carbon Dioxide Capture from Existing Coal-Fired plants study [1], except that the costs analysis in this report will be expressed in December 2009 U.S. dollars. This approach will enable comparison of this study results with the appropriately escalated results of the 2007 DOE/NETL study [1].

The capital cost estimates in this study will be assessed on a Total Investment Cost (TIC) level[d], and will be presented on an engineering, procurement, and construction (EPC) reimbursable basis with process and project contingencies. All costs will be estimated in December 2009 U.S. dollars. These costs will include all required equipment to complete the retrofit such as the new membrane based CO_2 capture system, the new CO_2 compression, and dehydration system, additional new balance of plant systems and modifications to the existing plant equipment and systems as required to support operation of the retrofitted plant.

Operating and maintenance (O&M) costs will be calculated for all systems. The O&M costs for the Base Case (pre-retrofit Conesville #5 Unit) will be based on the 2007 DOE/NETL study [1]. For the retrofit CO_2 capture system evaluations, additional O&M costs will be calculated for the new equipment. The variable operating and maintenance (VOM) costs for the new equipment included such categories as chemicals and desiccants, waste handling, maintenance material and labor, contracted services, and make-up power cost (MUPC) from the reduction in net electricity production. The fixed operating and maintenance (FOM) costs for the new equipment includes operating labor only.

Cost Estimation Basis

The following assumptions will be made in developing cost estimates for retrofit case:

1. December 2009 U.S. dollars
2. Outdoor installation
3. Investment in new utility systems (outside of the plant boundary [13]) is outside the scope
4. CO_2 product pipeline is outside the scope
5. No special limitations for transportation of large equipment

[d] The TIC cost documented in the reference study is consistent with the Total Plant Cost (TPC) nomenclature utilized by DOE/NETL.

6. No protection against unusual airborne contaminants (dust, salt, etc.)
7. No unusual wind storms
8. No earthquakes
9. No piling required
10. All releases can go to atmosphere – no flare provided
11. CO_2 Product compressor designed to API standards, all other pumps conform to ANSI
12. All heat exchangers designed to TEMA "C"
13. All vessels are designed to ASME Section VIII, Div 1.
14. Annual operating time is 7,446 hr/yr (85% capacity factor)
15. The investment cost estimate was developed as a factored estimate based on in-house data for the major equipment. Such an estimate can be expected to have accuracy of +/- 30%.
16. Process and project contingency will be added to the EPC to derive the TIC.
17. Make-up power cost was assessed at a 20-year levelized rate of 6.40 ¢/kWh (equivalent to a new Subcritical Pulverized Coal (Greenfield) Plant without carbon capture)
    Note: This is relevant to the EPRI Economics and not the capital cost development.
18. No purchases of utilities or charges for shutdown time have been charged against the project

Other exclusions from the cost estimate are as follows:

1. Soil investigation
2. Environmental permits
3. Disposal of hazardous or toxic waste
4. Disposal of existing materials
5. Custom's and Import duties
6. Sales/use tax
7. Forward escalation
8. Capital spare parts
9. Chemical loading facilities
10. Buildings except for compressor building and electrical substation
11. Financing cost
12. Owners cost
13. Guards during construction

14. Site medical and ambulance service
15. Cost & fees of authorities
16. Overhead high voltage feed lines
17. Cost to run a natural gas pipeline to the plant
18. Excessive piling
19. Contingency and risk

The costs used for consumption of fuel and chemicals in this project are shown in Table 3-1. The cost values are presumed to be July 2006 US dollars will be escalated to December 2009.

Table 3-1
Consumables Prices

Consumables	Units	Value
MEA	$/lb	0.95
Soda Ash	$/lb	0.26
Corrosion Inhibitor	$/lb	3.00
Activated carbon	$/lb	1.00
Molecular Sieve	$/lb	2.00
Diatomaceous Earth	$/lb	1.25
Coal	$/10^6 Btu	1.90
Natural Gas	$/10^6 Btu	7.12

The project and process contingencies applied to the capital expenditures are shown in Table 3-2.

Table 3-2
Project and Process Contingencies

Consumables	Project Contingency*	Process Contingency*
CO_2 Separation and Compression system	25%	18%
Flue gas Desulfurization (FGD)	11%	0%

* Percent of bare erected cost (i.e. subtotal direct cost in the investment tables for each case)

Capital costs (Table 3-3) and O&M costs (Table 3-4) for Case 1 (90% capture) for 2007 DOE/NETL study [1] will be utilized as a basis for comparison with this study results.

Table 3-3
2007 DOE/NETL Study Case 1 CO_2 Separation and Compression System Investment Costs

Acc't Code	Description	Pieces	Direct Manhours	Labor ($1,000)	Material ($1,000)	Total ($1,000)	%
11000	Heaters						0.00%
11200	Exchangers & Aircoolers		25,200	466	19,049	19,515	5.17%
12000	Vessel / Filters		6,638	123	5,018	5,141	1.36%
12100	Towers / Internals		29,859	552	22,571	23,123	6.12%
12200	Reactors						0.00%
13000	Tanks						0.00%
14100	Pumps		4,431	82	3,350	3,432	0.91%
14200	Compressors		60,663	1,122	45,856	46,978	12.43%
18000	Special Equipment		5,070	94	3,833	3,926	1.04%
	Sub-Total Equipment	**140**	**131,862**	**2,439**	**99,676**	**102,115**	**27.03%**
21000	Civil		175,815	3,253	6,977	10,230	2.71%
21100	Site Preparation						0.00%
22000	Structures		46,152	854	4,087	4,941	1.31%
23000	Buildings		24,175	447	1,196	1,643	0.43%
30000	Piping		362,619	6,708	17,942	24,650	6.52%
40000	Electrical		186,804	3,456	7,974	11,430	3.03%
50000	Instruments		153,839	2,846	12,460	15,306	4.05%
61100	Insulation		131,862	2,439	5,183	7,623	2.02%
61200	Fireproofing		65,931	1,220	1,495	2,715	0.72%
61300	Painting		32,965	610	698	1,308	0.35%
	Sub-Total Commodities		**1,180,161**	**21,833**	**58,011**	**79,844**	**21.13%**
70000	Construction Indirects					35,228	9.32%
	Sub-Total Direct Cost (Bare Erected Cost)		**1,312,023**	**24,272**	**157,687**	**217,188**	**57.48%**
71000	Construction Management					2,000	0.53%
80000	Home Office Engineering					29,400	7.78%
80000	Basic Engineering					5,000	1.32%
95000	License Fee	Excluded					0.00%
19400	Vendor Reps					1,750	0.46%
19300	Spare parts					2,900	0.77%
80000	Training cost	Excluded					0.00%
80000	Commissioning	Excluded					0.00%
19200	Catalyst & Chemicals	Excluded					0.00%
97000	Freight					4,700	1.24%
96000	CGL / BAR Insurance						0.00%
91400	Escalation to July 2006 Dollars					7,200	1.91%
	Total Base Cost					**270,138**	**71.50%**
	Contractors Fee					14,300	3.78%
	Total (EPC):					**284,438**	**75.28%**
93000	Project Contingency					54,297	14.37%
93000	Process Contingency					39,094	10.35%
	Total Investment Cost (TIC):					**377,829**	**100.00%**

Table 3-4
2007 DOE/NETL Study Case 1 CO_2 Separation and Compression System Operating & Maintenance Costs

Operating & Maintenance Costs	Subtotal ($1000/yr)	Total ($1000/yr)
Fixed O&M Costs		2,494
Operating Labor	2,494	
Variable O&M Costs		17,645
Chemicals	10,161	
Waste Handling & Contracted Services	767	
Maintenance (Materials and Labor)	6,716	
Feedstock O&M Costs		653
Natural Gas	653	
Levelized, Make-up Power Cost		62,194
Levelized, Make-up Power Cost (@ $6.40 ¢/kWh)	62,194	

4 REFERENCES

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