

# **INDUSTRIAL DESIGN AND OPTIMIZATION OF CO<sub>2</sub> CAPTURE, DEHYDRATION, AND COMPRESSION FACILITIES**

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## **ABSTRACT**

On an industrial scale process, a comprehensive engineering design and optimization study was conducted for CO<sub>2</sub> capture, dehydration, and compression facilities based on flue gases from natural gas and coal fired power plants. The HTC designer solvent was utilized in this chemical absorption process to achieve CO<sub>2</sub> recovery targets from 80 to 90%. The captured and conditioned CO<sub>2</sub>, with more than 99 mol% purity, was compressed to 150 barg and sent out at the boundary limit for enhanced oil recovery applications. The main design and engineering factors affecting the CO<sub>2</sub> capture, dehydration, and compression processes have been highlighted in this paper. The study provides a feasible engineering design and acceptable production cost taking into consideration all the technical, economic, and plant location factors. The study shows that it is advantageous to use the HTC designer solvent over the conventional monoethanolamine (MEA) solvent mainly due to its lower steam consumption, solvent losses, circulation rate, and cooling water requirements. Based on the objective function, the assumed industrial constraints, and the plant location factor, the production cost is estimated to be about 49 US\$/ton CO<sub>2</sub> for the 90% CO<sub>2</sub> recovery of 4.0 mol% CO<sub>2</sub> content in the flue gas of a natural gas combined cycle power plant. However, a substantial reduction in the production cost was reported for higher CO<sub>2</sub> contents in the flue gas of a coal power plant. For a similar CO<sub>2</sub> production capacity of 3307 ton per day from a 12 mol% CO<sub>2</sub> content in flue gas of a coal fired power plant, the production cost is about US\$ 30/ton CO<sub>2</sub>. This substantial reduction in the production cost is mainly because of the higher CO<sub>2</sub> contents in the flue gas.

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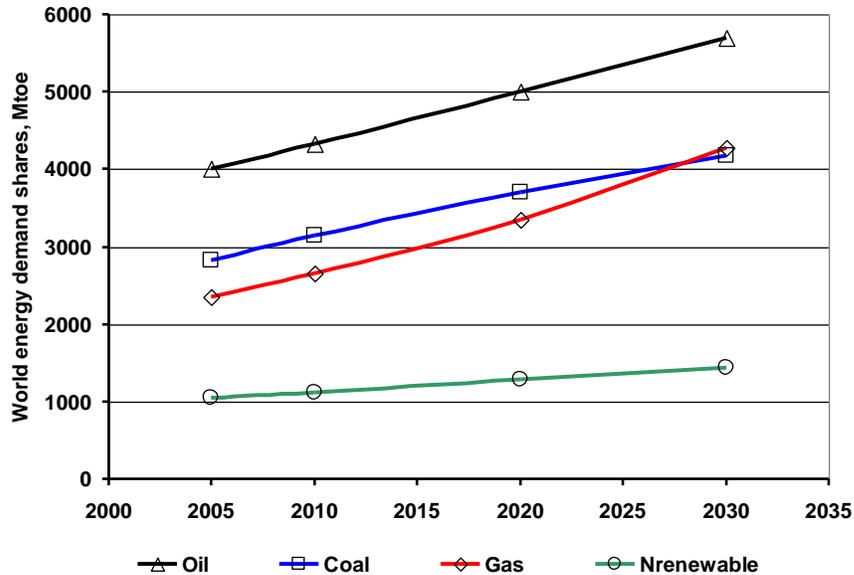
## INTRODUCTION

The removal of carbon dioxide, CO<sub>2</sub>, from gaseous mixtures by means of liquid absorbents will continue to be one of the leading industrial applications in the field of gas absorption for many decades to come from both industrial and environmental points of view. The removal of CO<sub>2</sub> by liquid absorbents is widely implemented in the field of gas processing, chemical production, and coal gasification. Today, many power plants also start considering the post-combustion option to meet the environmental regulation and to produce CO<sub>2</sub> for enhanced oil recovery applications. Some major industrial processes that require CO<sub>2</sub> removal to achieve specific cleanup targets are presented in Table I. As can be seen from this table, the cleanup target, which is the allowable extent of CO<sub>2</sub> in the treated gas to meet product specifications or environmental regulations, varies from process to process. For example, the cleanup target for CO<sub>2</sub> is 1% by volume for pipeline gas. However, for ammonia and LNG manufacture, CO<sub>2</sub> must be reduced to 16 ppm and 50 ppm, respectively. For power plant flue gases, it is acceptable to set the cleanup target to less than 0.5% for NGCC Power Plants and less than 1.5% for Coal Power Plants, from an environmental and economical points of view.

From an environmental point of view, Figure 1 illustrates actual data of fuel consumption in 2005 and an estimation of energy demand of different fuel types for years to come, from 2010 to 2030. It can be seen that the world energy demand will gradually increase at rates of 10 – 15 % every 10 years. This increase will raise the CO<sub>2</sub> emissions without doubt to about 50% by 2030 in comparison with the current level of CO<sub>2</sub> emissions. The industrial countries (North America, Western Europe and OECD Pacific) contribute in this jump of emission by 70% comparing to the rest of World and about more than 60% of these emissions will come from power generation and industrial sectors [1].

**Table I - CO<sub>2</sub> removal application in major industrial processes**

| <i><b>Process</b></i>                             | <i><b>Common cleanup targets</b></i> |
|---|--------------------------------------|
| Hydrogen Manufacture                              | < 0.1% CO <sub>2</sub>               |
| Ammonia Manufacture                               | < 16 ppm CO <sub>2</sub>             |
| Natural Gas Purification:                         |                                      |
| Pipeline Gas                                      | < 1% CO <sub>2</sub>                 |
| LNG Feedstock                                     | < 50 ppm CO <sub>2</sub>             |
| Synthesis Gas for Chemical (H <sub>2</sub> /CO)   | < 500 ppm CO <sub>2</sub>            |
| Coal Gasification                                 | ~ 500 ppm CO <sub>2</sub>            |
| Ethylene Manufacture (steam cracker gas treating) | ~ 1 ppm CO <sub>2</sub>              |
| Power Plants:                                     |                                      |
| NGCC Power Plant                                  | < 0.5 % CO <sub>2</sub>              |
| Coal Fired Power Plant                            | < 1.5% CO <sub>2</sub>               |



**Figure 1 - World Energy Demand Shares by Fuel Types [1]**

Greenhouse gas emissions, including CO<sub>2</sub>, should be limited, as recommended at the Kyoto Conference, Japan, in December 1997, and Bali Conference, Indonesia, in December 2007 [2]. Despite the strong recommendations and plans in these conferences, there are hardly any investments in the CO<sub>2</sub> capture facilities by the industrial sectors toward meeting these cleanup targets mainly because of the high production cost of CO<sub>2</sub> from flue gases. One of the methods to reduce the CO<sub>2</sub> capture cost is to design an energy efficient gas absorption process. Based on the findings of a recent conceptual engineering study, HTC Purenergy estimated the production cost to be US\$ 49/ ton CO<sub>2</sub> (US\$ 54/ tonne CO<sub>2</sub>) for the 90% CO<sub>2</sub> recovery of 4 mol% CO<sub>2</sub> content in the flue gas of NGCC, as documented elsewhere [3]. In this work, a substantial reduction in the production cost was reported for higher CO<sub>2</sub> contents in the flue gas of coal power plant.

## SCOPE OF WORK AND CONSTRAINTS

The overall scope of work for this study is to provide an energy efficient CO<sub>2</sub> capture plant design from a coal power plant to produce 3307 ton per day. The plant shall include flue gas pre-conditioning and post-capture conditioning including compression.

The design of the facility is based on the specified flue gas conditions, CO<sub>2</sub> product specifications, and constrains. The flue gas conditions utilized in the design is presented in Table II. The bulk removal of CO<sub>2</sub> is recommended to capture 90% of the CO<sub>2</sub> from the flue gas of this coal fired power plant.

The main CO<sub>2</sub> product specifications are specified as follows:

- Recommended design level of oxygen content in the product stream is 50 ppm.
- Target for water content is < 10 ppm and -50°F dew point. This specification is required after the compression facilities and before the pipeline transfers the product to the client.
- Maximum nitrogen content in the CO<sub>2</sub> product is 2%.
- Minimum CO<sub>2</sub> product pressure at the compressor inlet is 3 psig.

**Table II - Flue Gas Specifications from a coal power plant.**

| <b>Item</b>      | <b>Value</b>    |
|------------------|-----------------|
| Flue gas rate    | 528 MMSCFD      |
| Temperature      | 180°F           |
| Pressure         | 14.145 psia     |
| N <sub>2</sub>   | 63.47 mol%      |
| CO <sub>2</sub>  | 12 mol%         |
| H <sub>2</sub> O | 18.5 mol%       |
| O <sub>2</sub>   | 6 mol%          |
| SO <sub>2</sub>  | 120 ppm         |
| NO               | 179 ppm         |
| NO <sub>2</sub>  | 7 ppm           |
| Particulate      | < 0.02 gr / SCF |

## **PLANT DESIGN**

The technology utilized by HTC Purenergy is based on the bulk removal of CO<sub>2</sub> by liquid chemical absorbents from flue gas streams. This is a suitable process technique for treating high-volume gas streams containing carbon dioxide at low pressure produced from utility power plants. The formulated solvent recommended for the CO<sub>2</sub> absorption process is a mixed amine solvent consisting mainly of primary, secondary, and tertiary amines, which are available commercially. The mixture concentration and the amine ratios are proprietary information.

### **Inputs and Constraints:**

The main design and operation constraints of the plant can be presented as follows:

- Flow rate of the flue gas (slipstream) to design the plant is 456 lb/s (207 kg/s) at 180°F and 14.145 psia. This flow rate is calculated to produce 3307 ton per day (3000 metric tonne per day) of CO<sub>2</sub> at a 90% CO<sub>2</sub> recovery.
- Minimum CO<sub>2</sub> product pressure at the compressor inlet is 3 psig.
- Low-pressure steam for reboiler is assumed to be available at 302°F (150°C) and 58 psia (4 bar). The return pressure of the condensate is assumed at 102 psia (7 bar).
- Low-pressure steam for reclaimer is assumed to be available at 392°F (200°C) and 102 psia (7 bar). The return condensates can be routed to a low-pressure condensate collection system.
- Cooling water is assumed to be available at 80°F and maximum allowable return temperature is 105°F.
- Plant is designed to meet product specification as outlined in the scope of work.
- Plant is designed to minimize harmful emissions to air.
- Plant is designed to produce minimum liquid and solid discharges.

Using the ProMax® process simulation software from Bryan Research & Engineering, all of the CO<sub>2</sub> Capture equipment units are designed for 90% CO<sub>2</sub> recovery using a mixed amine solvent

with additional capacity in the pumps and heat exchanger areas. This design provides excess capacity in order to accommodate any future utilization of new solvents, new packing, new operating conditions, change in the CO<sub>2</sub> content, new cleanup targets, and/or any new optimization parameters.

In addition, the following factors were considered in the design:

- Pump overall efficiency: 65%
- Pump pressure differential, delivery pressure, or dynamic head are assumed
- Number of ideal stages in the SO<sub>2</sub>/cooler section in the absorption column: 3
- Number of ideal stages in the solvent section in the absorption column: 3
- Number of ideal stages in the washing section in the absorption column: 2
- Number of ideal stages in the stripping column including reflux and reboiler: 10
- The height of each stage: 6 to 9 ft depending on the packing type.

### **Process Description and Process Flow Diagram**

Figure 2 presents a simplified process flow diagram representing the configuration of the proposed CO<sub>2</sub> Capture Plant. Inlet Gas Blower (K-101) is designed to boost the flue gas from atmospheric pressure to about 1.5 psig. This is required to overcome the pressure drop of the absorber column and at the same time to maintain the back pressure on the power plant equipment.

The absorber column (C-101) is comprised of an inlet gas SO<sub>2</sub>/cooling section, absorption section, the off-gas washing section, and a flue gas vent stack. The top two sections will be packed by random packing or high efficiency structured packing. The purpose of the off-gas washing section is to recover solvent entrained in water vapors from the absorber section and to cool the off-gas at a temperature to help maintain a water balance across the plant. Each section is separated by a chimney tray that allows gas to pass up through to the next section but ensures liquid separation.

Flue gas from the blower enters the bottom inlet gas scrubber section of C-101 below stage 1. Water/Chemical drawn from the bottom of the column is circulated via the pump (P-101) through the flue gas water cooler (E-101) and back to the column where it is distributed above the third stage. The flue gas in contact with the scrubber water is cooled down before entering the absorber section through the chimney tray. The blowdown/Bleed stream is transferred for further treatment and utilization. The SO<sub>2</sub> treatment process selected for this project is the Ammonium Sulfate FGD system, which has the capability to reduce flue gas SO<sub>2</sub> concentrations to below 10 ppm. Ammonia reagent will be required for the process. The process can utilize anhydrous ammonia, aqueous ammonia or a waste ammonia source. The conceptual system has been developed to produce a liquid ammonium sulfate product which can be marketed for direct application or further processing.

As the feed gas passes up from the bottom of the absorber section, lean amine enters the top of the absorber solvent section and is distributed over the packing material. The lean amine in counter flow comes in contact with the gas and absorbs the CO<sub>2</sub> out of the flue gas stream leaving the remaining gas to pass through the top chimney tray and into the absorber off-gas wash section. In order to avoid significant solvent losses, the top wash section is used to cool the off-gas and recover entrained chemical solvent as well as condense water vapors to maintain a plant water balance. The wash water is circulated from the wash section via pump P-106 through exchanger E-106 and back to the top wash section.

The CO<sub>2</sub> rich amine is collected at the bottom of the absorber section and pumped via P-102 through a lean/rich exchanger (E-102) to near the top of the regenerator column. This column is also

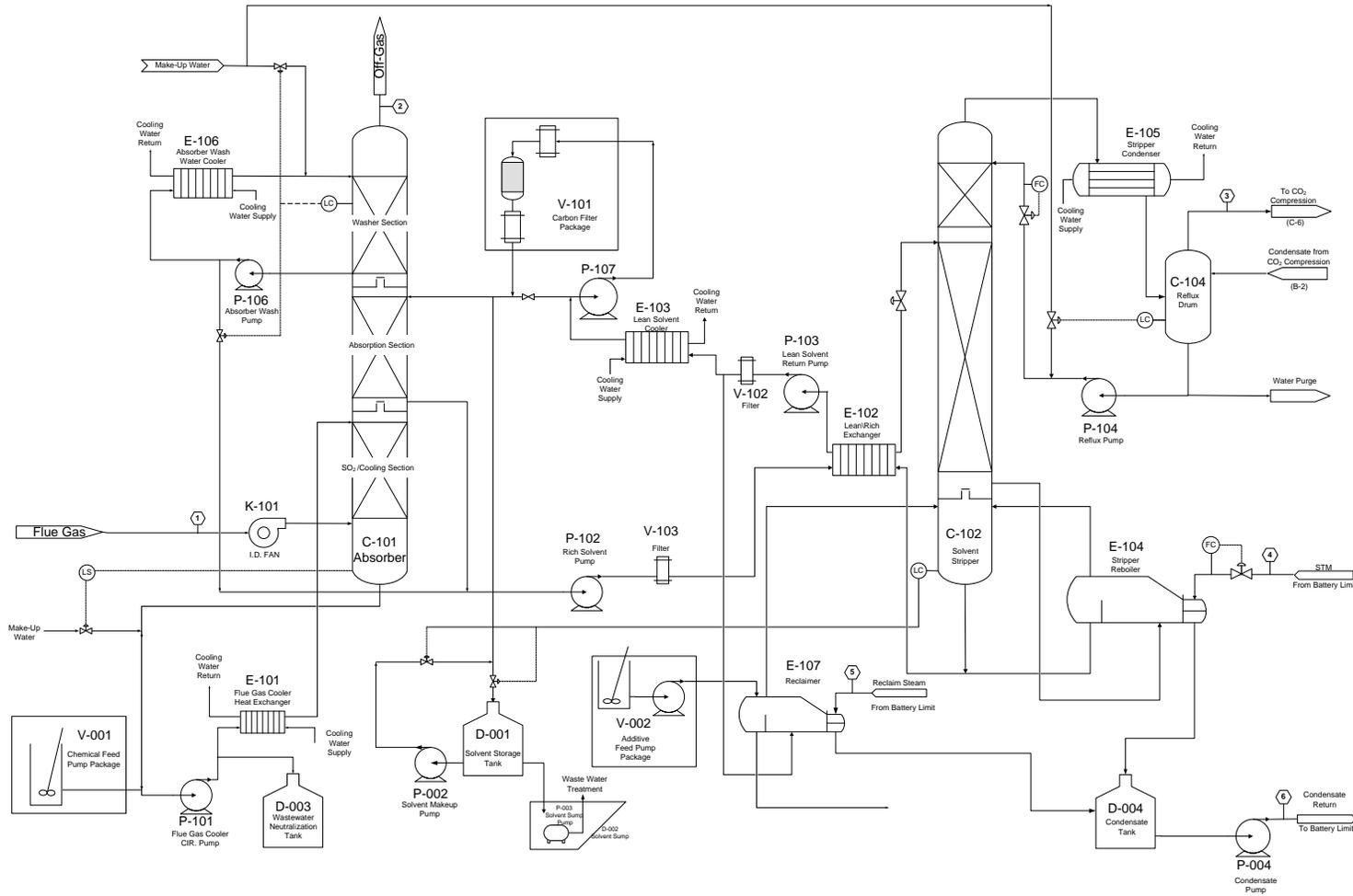
packed with random or high efficiency structured packing. The reboiler (E-104) located near the bottom and connected to the stripper provides the heat necessary to strip out the CO<sub>2</sub> from the chemical solvent. Stripping steam produced from the reboiler passes upward through the stripper column. As the rich amine descends in counter flow, CO<sub>2</sub> is stripped out of the chemical solvent, in effect, reversing the solvent absorption reaction. The stripped lean amine accumulates in the bottom of the stripper column and reboiler where it exits to the lean/rich exchanger in order to utilize the lean amine heat from the regeneration process to preheat the rich amine prior to entering the stripper column, and at the same time, reduce the cooling load on the lean amine.

From the top of the stripper column, CO<sub>2</sub> product, water vapor, and entrained amine enter the reflux condenser (E-105) where the majority of the liquid is condensed, accumulated in the reflux drum (C-104) and pumped back to the top of the stripper column reflux section via pump P-104 with excess sent back to the absorber section as makeup water. The final CO<sub>2</sub> product is delivered to the compression section where it is dried and compressed through a multi-stage compressor.

A reclaimer (E-107) is used for regenerating the chemical solvent that is degrading due to a buildup of heat stable salts (HSS) and other impurities in the amine solution. Heat and caustic solution injection are the prime process ingredients for regeneration of the solvent. With a slip stream of less than 2% lean amine into the reclaimer delivered by P-103, as illustrated in the process flow diagram, the introduction of heat and caustic releases the solvent in the vapor phase from the degradation product back into the stripper, leaving behind a non-regenerable sludge at the bottom of the reclaimer, which can be collected, measured, analyzed, and disposed of.

From the lean/rich exchanger the lean amine is pumped via P-103 to the lean amine cooler E-103 where it is cooled down before entering back into the absorber column. A 5 to 10% slipstream of lean amine is passed through a carbon filter package unit (V-101) to remove degradation products prior to returning to the absorber column. V-101 is comprised of a mechanical cartridge filter to remove rust and other debris followed by a carbon drum containing activated carbon and finally another mechanical cartridge filter to remove entrained carbon in the stream. A solvent surge/storage tank (D-001) and solvent charge pump (P-002) are used for solvent makeup to the system, provide storage, and provide surge capacity to ensure system volume is maintained.

To meet the CO<sub>2</sub> pipeline specification in terms of pressure and water content, a TEG dehydration process was integrated within the compression facility. The CO<sub>2</sub> product stream is compressed from 1.7 bar to 150 bar and air cooling is utilized to cool the CO<sub>2</sub> stream between compressor stages. To meet the water specification of about 4 lb/MMSCF water content, it is required to use a dehydration process. It is recommended to install the dehydration system between compressor stages at pressure > 520 psia. This will reduce the size of the equipment and to make the dehydration process more efficient. If a water content of less than 4 lb/MMSCF is required, other options for dehydration can be implemented by using a stand alone molecular sieve system or TEG/molecular sieve combined system. The first option requires larger adsorption beds. In the second option, the TEG system is followed by a smaller diameter molecular sieve in order to meet the cleanup target of < 10 ppm of water content. The costs of these two options are comparable.



**Figure 2 - HTC Simplified Process Flow Diagram for the CO<sub>2</sub> Capture Plant.**

## Environment

The CO<sub>2</sub> capture plant was designed to produce minimum liquid and solid discharge, as well as, to minimize harmful emissions to air.

### Air Emissions

Air emissions primarily originate from the absorber tower and the TEG dehydration unit, if used. These emissions will primarily consist of gases and contaminants originating from the power plant and amine system solvent evaporation and carryover.

NO<sub>x</sub> is comprised of about 95% NO and 5% NO<sub>2</sub>. NO<sub>2</sub> is water soluble and is removed in the absorber column. NO as expected is passed through the column without being removed by the solvent with a rate of 10.38 lbmol/h, which is the same amount originally in the flue gas.

Flue gas from coal fuel is inherently high in SO<sub>2</sub>, and as a result pre-treatment is necessary to remove the majority of the SO<sub>2</sub> to ensure that the solvent does not degrade. Any SO<sub>2</sub> entering the absorber will react with amine to form a sulfate salt. In this reaction, some amine is consumed, and the reaction product is removed by the reclaimers as sludge.

All amines are manufactured from ammonia. Although it is a stable product, some ammonia dissociation is possible. Table III shows ammonia in the absorber exit gas based on equilibrium calculations. Dissociation of amine to ammonia may occur when the amine is degraded. If the amine is properly maintained, its degradation rate should be small and its impact on ammonia generation should be small. In the off gas, the ammonia quantified at a rate of 0.178 lbmol/h (3 lb/h) and concentration of about 4 ppm., which is less than the 25 ppm reported from conventional amine processes. From this table, it can also be seen that the solvent loss due to carryover and evaporation is 0.028 lbmol/h (1.7 lb/hr) and 0.618 ppm. The rest of the emissions are mainly nitrogen about 82%, oxygen about 7.8%, and water vapor about 8.6%. The emissions to air from the TEG and molecular sieve dehydration processes are negligible.

**Table III - Absorber Off Gas Constituents**

| item         | Off Gas          |               |            |
|--------------|------------------|---------------|------------|
|              | lbmol/h          | mol%          | ppm        |
| N2           | 36805.797        | 82.11963%     | 821196.256 |
| CO2          | 648.651          | 1.44724%      | 14472.443  |
| H2O          | 3874.553         | 8.64475%      | 86447.478  |
| O2           | 3479.325         | 7.76293%      | 77629.309  |
| SO2          | 0.416            | 0.00093%      | 9.290      |
| NO           | 10.380           | 0.02316%      | 231.590    |
| NO2          | 0.406            | 0.00091%      | 9.052      |
| NH3          | 0.178            | 0.00040%      | 3.964      |
| HTC Solvent  | 0.028            | 0.00006%      | 0.618      |
| <b>Total</b> | <b>44819.732</b> | <b>100.0%</b> |            |

### Liquid waste

Liquid amine solvents of any concentration are not to be disposed of at the sewer or any bodies of water. All internal condensates from the reflux drum and compressor intercoolers and after coolers are recycled to the solvent system as the makeup water for the solvent dilution. This reduces the raw

water make up requirement for the plant. This water is sent to a holding neutralization tank where pH adjustment is made before recycle.

### **Solid Discharges**

The solid wastes from the CO<sub>2</sub> capture plant are:

- Spent carbon bed – this is an “exempt” waste that can be disposed of in land fills or recycled.
- Sludge from reclaimer – these are heat stable salts such as sulfates, chlorides, nitrates as well as metal oxides and other degradation/corrosion products. Similar to standard amine systems, the sludge stream requires disposal in an industrial waste disposal facility.
- The carbon filter is regenerated by steam once or twice and then disposed of in a landfill.
- Mechanical filters are reusable types and do not require routine disposal.

### **Production Cost.**

The indicative estimate for the direct cost was prepared based on price information received from similar projects. This is an order of magnitude estimate with an accuracy range of  $\pm 40\%$ . The total estimated capital cost is about US\$ 165 million. This indicative cost includes equipment cost (SO<sub>2</sub> treatment unit, CO<sub>2</sub> capture facility, dehydration unit, and compression facility), home office, construction management, commissioning and start up costs, raw material and solvent initial fill, contingency and risk allowance. The total annual operating cost is estimated at about US\$ 25 million. The majority of the operating costs are based on the consumption of utilities including steam, electricity, and cooling water. Other costs include chemical consumption, insurance and taxes, and labour associated with the operation and maintenance along with overhead. With the use of the formulated solvent, steam costs represent 40% of the total operating costs compared to conventional MEA steam costs of 70-80% of total operating costs. The other two large items in the operating cost are the electricity, representing 20% of the operating cost, and the maintenance cost, representing 15% of the operating cost. The cooling water costs represent less than 5% of the cost depending on the source of cooling.

Based on the capital cost, operating cost, and the required production capacity, the estimated production cost is US\$ 30/ton CO<sub>2</sub> (US\$ 33/metric tonne CO<sub>2</sub>). This cost includes the cost of the SO<sub>2</sub> treatment, CO<sub>2</sub> capture, dehydration, and compression costs. These amounts may be substantially reduced if Government/Industry contributions are made to the project, which effectively reduce the CAPEX costs and/or subsidizing the utilities cost such as the steam and electricity.

### **Results and Conclusions**

Based on these inputs, constrains, cleanup targets, and assumptions; the main design findings are:

- Diameter of the absorber column is 32.8 ft.
- Diameter of the regenerator column is 19.7 ft.
- Columns heights are ranging from 100 to 130 ft depending on the packing type used.
- Plant estimated layout is about 250ft x 250ft area.

The energy efficiency of a CO<sub>2</sub> capture plant depends primarily on the performance of the solvent and how well the plant configuration (flow sheet) is optimized. In earlier flue gas amine plant designs, MEA was the primary solvent used and its concentrations were limited to 20 wt% to minimize equipment corrosion. However, this lower concentration has resulted in higher steam consumption because steam is needed to boil off the diluting water in the reboiler. Recent development in controlling corrosion and degradation has allowed the industry to increase the solvent concentration to about 30 wt% thus decreasing the steam demand. The use of a split flow configuration has an impact in lowering the steam consumption. A recent DOE study [4] shows the steam consumption for an existing CO<sub>2</sub> plant using 18 wt% MEA (Kerr McGee Process) is 3.45 lb of steam per lb of CO<sub>2</sub> for amine regeneration. A modern process that uses 30 wt% MEA is expected to use 1.67 lb of steam per lb of CO<sub>2</sub> using a split flow configuration for amine regeneration.

The HTC formulated solvent has a lower steam usage than the conventional MEA solvent. Based on the material and energy balances for the plant designed in this study, the reboiler steam rate is about 406,800 lb/h at 302°F and 43.5 psig by using the proposed formulated solvent on a standard flow sheet, without implementing any split flow configurations. Based on this the reboiler steam consumption is estimated at about 1.47 lb steam/ lb CO<sub>2</sub>, which is much less than the reported steam for MEA solvent.

The Nitrogen content in the product stream is about 116 ppm, which is less than the specified spec of 2%. The oxygen content is about 20.4 ppm, which is less than the specified spec of 50 ppm in the scope of work. The water content before the compression is about 2.8 mol%, which is within the industrial range for the amine capture process type. A substantial amount of this water content will be removed within the first few stages of compression using knock-out drums during the cooling process between the compression stages. At about 520 psig pressure and 120°F, within the 8-stage compression/cooling system, the CO<sub>2</sub> product will be routed to a TEG dehydration process which will reduce the water content to about 4 lb per MMSCF. This is sufficient water content to prevent hydrate formation and corrosion during the transportation of CO<sub>2</sub> product at about 2,700 psig. The target of less than 10 ppm can be achieved by using a stand alone molecular sieve system or TEG/molecular sieve combined system. The first option requires larger adsorption beds. In the second option, the TEG system is followed by smaller diameter molecular sieves in order to meet the cleanup target of < 10 ppm of water content. The costs of these two options are comparable.

Table IV presents the main findings for CO<sub>2</sub> capture from the referenced coal plant and a NGCC power plant, each designed to produce about 3307 ton per day (3,000 TPD metric).. To produce the same capacity of CO<sub>2</sub>, it can be seen that only one train with smaller column diameters is required in case of the coal power plant and two trains with larger column diameters are required in case of the NGCC Power Plant. This is mainly due to processing a larger flue gas with lower CO<sub>2</sub> content in the NGCC power plant than the coal power plant case. Subsequently, a substantial reduction in the capital and production cost was reported for higher CO<sub>2</sub> content in the flue gas of the coal power plant.

Based on the capital cost, operating cost, and the required production capacity, the estimated production cost is about 49 US\$/ton CO<sub>2</sub> of 4.0 mol% CO<sub>2</sub> content in the flue gas of NGCC power plant. For a similar CO<sub>2</sub> production capacity of 3307 ton per day from a 12 mol% CO<sub>2</sub> content in flue gas of a coal fired power plant, the production cost is estimated at about US\$ 30/ton CO<sub>2</sub>. This substantial reduction in the production cost is mainly because of the higher CO<sub>2</sub> contents in the flue gas of the coal power plant.

**Table IV - CO<sub>2</sub> Capture plant parameters to produce 3307 ton per day from flue gas of two different Sources, Coal Power Plant and NGCC Power Plant.**

| <b>Parameter</b>                                 | <b>Coal Power Plant,<br/>This work</b> | <b>NGCC Power Plant [3]</b> |
|--|--|-----------------------------|
| CO <sub>2</sub> Production Capacity, ton per day | 3307                                   | 3307                        |
| CO <sub>2</sub> recovery, %                      | 90                                     | 90                          |
| CO <sub>2</sub> concentration in flue gas, mol%  | 12                                     | 4                           |
| Number of trains                                 | 1                                      | 2                           |
| Flue gas rate,                                   | 528 MMSCFD                             | 920 MMSCFD                  |
| Absorber diameter, ft                            | 32.8                                   | 39.4                        |
| Regenerator diameter, ft                         | 19.7                                   | 19.7                        |
| Capital Cost, million US\$                       | 165                                    | 227                         |
| Operating cost, million US\$                     | 25                                     | 51                          |
| Production cost, US\$/ton                        | 30                                     | 49                          |

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