

**Bench-Scale Development of a Hot Carbonate Absorption Process with  
Crystallization-Enabled High Pressure Stripping for Post-Combustion CO<sub>2</sub> Capture**

**–Preliminary Techno-Economic Study Results and Methodology**

**DOE Award Number: DE-FE0004360  
Recipient: Illinois State Geological Survey  
Prairie Research Institute  
University of Illinois at Urbana-Champaign  
Tel: 217-244-4985; Fax: 217-333-8566**

**Principal Investigator: Yongqi Lu**

**Milestone Deliverable by:  
Carbon Capture Scientific, LLC  
4000 Brownsville Road, South Park, PA 15129  
Tel: 412-854-6713; Fax: 412-854-6610**

**Submitted to**

**U. S. Department of Energy  
National Energy Technology Laboratory**

**Project Manager: Andrew Jones  
Email: [andrew.jones@netl.doe.gov](mailto:andrew.jones@netl.doe.gov)**

**January 29, 2014**

## **Disclaimer**

This report was prepared as an account of work sponsored by an agency of the United States Government. Neither the United States Government nor any agency thereof, nor any of their employees, makes any warranty, express or implied, or assumes any legal liability or responsibility for the accuracy, completeness, or usefulness of any information, apparatus, product, or process disclosed, or represents that its use would not infringe privately owned rights. Reference herein to any specific commercial product, process, or service by trade name, trademark, manufacturer, or otherwise does not necessarily constitute or imply its endorsement, recommendation, or favoring by the United States Government or any agency thereof. The views and opinions of authors expressed herein do not necessarily state or reflect those of the United States Government or any agency thereof.

## Abstract

This report summarizes the methodology and preliminary results of a techno-economic analysis (TEA) on a hot carbonate absorption process (Hot-CAP) with crystallization-enabled high pressure stripping for post-combustion CO<sub>2</sub> capture (PCC). This analysis was based on the Hot-CAP that is fully-integrated with a 550 MWe subcritical steam cycle, pulverized coal-fired power plant adopted in Case 10 of the DOE/NETL's Cost and Performance Baseline for Fossil Energy Plants.

The TEA addresses several important aspects of the Hot-CAP for PCC application, including process design and simulation, equipment sizing, technical risk and mitigation strategy, performance evaluation, and cost analysis. Results show that the net power produced in the subcritical power plant equipped with the Hot-CAP is 611 MWe, greater than that (550 MWe) with benchmark 30 wt% monoethanolamine (MEA). The total capital cost for the Hot-CAP, including CO<sub>2</sub> compression, is \$399 million, less than that for the MEA-based process (\$493 million). O&M costs for the power plant with Hot-CAP are \$176 million annually, less than that with benchmark MEA (\$178 million). The 20-year levelized cost of electricity (LCOE) for the power plant with Hot-CAP, including CO<sub>2</sub> transportation and storage, is 119.6 mills/kWh, a 59% increase over that for the plant without CO<sub>2</sub> capture. The LCOE increase caused by CO<sub>2</sub> capture for the Hot-CAP is 31% lower than the benchmark MEA process.

## Table of Contents

<b>Abstract</b> .....	iii
<b>List of Figures</b> .....	vi
<b>List of Tables</b> .....	vii
<b>Executive Summary</b> .....	viii
<b>1 Introduction</b> .....	1
<b>2 Design Basis</b> .....	2
2.1 Power Plant Design Criteria .....	2
2.1.1 General.....	2
2.1.2 Site-Related Conditions .....	2
2.1.3 Meteorological Data.....	2
2.1.4 Technical Assumptions and Data.....	3
2.1.5 Environmental/Emissions Requirements .....	5
2.2 PCC Design Criteria .....	5
2.2.1 General.....	5
2.2.2 Flue Gas Feed Specification .....	5
2.2.3 Design CO <sub>2</sub> Product Specifications .....	6
2.2.4 Utility Commodity Specifications .....	6
2.2.5 Process Water Streams.....	7
<b>3 Simulations and Design of a Hot Carbonate Absorption Process with Crystallization-Enabled High Pressure Stripping</b> .....	8
3.1 Overview and Description of Hot-CAP Process.....	8
3.2 Risk Analysis and Mitigation Strategy for the Hot CAP Process.....	9
3.2.1 Risk A Mitigation Strategy .....	10
3.2.2 Risk B Mitigation Strategy .....	13
3.2.3 Risk C Mitigation Strategy .....	14
3.2.4 Risk D Mitigation Strategy .....	14
3.2.5 Risk E Mitigation Strategy.....	14
3.3 Design of Hot-CAP process.....	15
3.3.1 Design of Absorption Column .....	15
3.3.2 Design of Stripping Column .....	16
3.3.3 Design of Crystallization Tanks.....	18

3.4 Mass and Heat Balance of Hot-CAP Process .....	20
<b>4 Cost Estimation .....</b>	<b>23</b>
4.1 Cost Estimation Methodology .....	23
4.1.1 Capital Cost.....	23
4.1.2 O&M Costs .....	25
4.2 Financial Modeling Basis .....	26
4.3 Performance Summary of Hot-CAP PCC .....	27
4.4 Capital Cost Estimate for Hot-CAP PCC .....	29
4.5 Performance Summary of Subcritical PC Plant with Hot-CAP PCC.....	35
4.6 Capital Cost Estimate for Subcritical PC Plant with Hot-CAP PCC.....	37
4.7 O&M Cost Estimate for Subcritical PC Plant with Hot-CAP PCC.....	39
4.8 LCOE Estimate for Subcritical PC Plant with Hot-CAP PCC .....	40
<b>5 Summary.....</b>	<b>42</b>
<b>References.....</b>	<b>43</b>
<b>Appendix - Acronyms and Abbreviations .....</b>	<b>44</b>

## List of Figures

Figure 3.1. Schematic diagram of the proposed Hot-CAP.....	9
Figure 3.2. Effect of temperature and CTB percentage on solubility of $K_2CO_3$ .....	9
Figure 3.3. Schematic of Hot-CAP technical risks: A,B,C,D,and E.....	9
Figure 3.4. Experimental results of $CO_2$ absorption performance in 40 wt% $K_2CO_3$ solution with various promoters.....	11
Figure 3.5 Simulation results with no promoter.....	12
Figure 3.6. Simulation results for 0.5 M PZ as a promoter.....	12
Figure 3.7. Simulation results for 1.0 M DEA as a promoter.....	13
Figure 3.8. Simulation results for 1.0 M PZ as a promoter.....	13
Figure 3.9. Multiple crystallizer design developed to address Risk D.....	15
Figure 3.10. Experimental VLE data for $CO_2$ in 60 wt% $K_2CO_3$ equivalent solution.....	16
Figure 3.11. Experimental VLE data for water vapor in 60 wt% $K_2CO_3$ equivalent solution.....	17
Figure 3.12. Schematic of crystallizer structure for Hot-CAP process.....	20
Figure 3.13. Schematic flowchart of Hot-CAP process with mass and heat balance.....	22

## List of Tables

Table 2.1. Illinois No. 6 Coal Properties.....	3
Table 2.2. PCC-independent PC auxiliary loads breakdowns.....	4
Table 2.3. Air Emissions targets.....	5
Table 2.4. Flue gas composition and CO <sub>2</sub> capture process operation condition.....	6
Table 3.1. Technical risks of Hot-CAP and mitigation strategies.....	10
Table 3.2 Estimation of volumes for five stages of crystallizers.....	19
Table 4.1 Cost Estimate Basis for Subcritical PC Plant with CO <sub>2</sub> Capture.....	24
Table 4.2. Hot CAP PCC capture section utilities.....	28
Table 4.3 Hot CAP PCC CO <sub>2</sub> compression section utilities.....	29
Table 4.4. Hot CAP-based CO <sub>2</sub> capture section major equipment list.....	30
Table 4.5. Hot CAP-based compression section major equipment list.....	31
Table 4.6. Hot CAP-based PCC CO <sub>2</sub> capture section total field cost.....	32
Table 4.7. Hot CAP-based PCC CO <sub>2</sub> compression section total field cost.....	33
Table 4.8. Hot CAP-based PCC total field cost.....	34
Table 4.9. Subcritical PC plant with Hot-CAP PCC performance summary.....	36
Table 4.10. Subcritical PC Plant with Hot CAP-Based PCC Capital Cost Estimate.....	38
Table 4.11. Subcritical PC Plant with Hot CAP-Based PCC O&M Costs.....	40
Table 4.12. Subcritical PC Plant with Hot CAP-Based PCC LCOE.....	41

## Executive Summary

A techno-economic analysis (TEA) was performed to assess a hot carbonate absorption process (Hot-CAP) with crystallization-enabled high pressure stripping which is fully integrated with a coal-fired power plant for post-combustion CO<sub>2</sub> capture (PCC). The TEA addresses several important aspects of the Hot-CAP development, including process design and simulation, equipment sizing, technical risk and mitigation strategy, performance evaluation, and cost analysis.

### Baseline Hot-CAP Design and Simulation

The TEA study was based on the Hot-CAP equipped with a subcritical steam cycle, pulverized coal-fired (PC) power plant adopted in Case 10 of the DOE/NETL's report on "Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity" (November 2010, Report No. DOE/NETL-2010/1397), which corresponds to a nominal 550 MWe (net) power plant that utilizes a benchmark 30 wt% MEA-based absorption system for CO<sub>2</sub> capture. In the current study, the Hot-CAP replaces the MEA-based system in the original case to achieve 90% CO<sub>2</sub> capture.

The Hot-CAP integrates three major unit operations, CO<sub>2</sub> absorption, potassium bicarbonate crystallization, and CO<sub>2</sub> stripping. A unique feature of the process is the CO<sub>2</sub> stripping at an elevated pressure enabled by the use of a bicarbonate slurry so that the energy usage for CO<sub>2</sub> stripping and compression can be reduced.

In the baseline Hot-CAP, the solvent is a 40 wt% K<sub>2</sub>CO<sub>3</sub> solution containing 0.5M piperazine (PZ) as a promoter with the CO<sub>2</sub> loading changing from ~20% (lean) to ~45% (rich) carbonate-to-bicarbonate (CTB) conversion in the absorber. The absorption operates at 60-67°C. The absorber consists of two parallel columns, each with effective packing of 13-m in height and 14.8-m in diameter.

A configuration of five consecutive stirred tanks is adopted for the bicarbonate crystallization. Each crystallizer employs submerged coils for heat exchange, a draft-tube for internal circulation of magma, and a downward-directed propeller agitator for liquid circulation control. The total volume of the crystallizers is estimated to be 3,220 m<sup>3</sup>, corresponding to a total residence time of ~15 minutes.

CO<sub>2</sub> stripping is performed at 6 bar and 181°C (reboiler temperature). A 60 wt% K<sub>2</sub>CO<sub>3</sub> slurry is employed and the CTB conversion of the slurry is reduced from 79% to 29% during the stripping. A single stripping column with effective packing of 10-m in height and 7.3-m in diameter can fulfill this purpose.

### Energy Use Performance

The Hot-CAP consumes less steam for CO<sub>2</sub> stripping than the benchmark MEA process. The steam extraction from the PC power plant equipped with the Hot-CAP includes 247 tonne/hr with 800 kPa pressure and 275 tonne/hr with 1,300 kPa pressure, which is equivalent to a total

heat requirement of 395.7 MW (thermal) or 2,382 kJ/kg CO<sub>2</sub> captured. In comparison, the MEA process in Case 10 uses 876 tonne/hr steam with 507 kPa pressure. In addition, the stripped CO<sub>2</sub> product gas from the Hot-CAP has a higher pressure than that from the MEA process (6 vs. 1.6 bar), resulting in a lower compression work requirement.

Overall, the net power produced in the PC power plant equipped with the Hot-CAP is 611 MWe, greater than that with the benchmark MEA process (550 MWe). Consequently, the thermal efficiency of the PC plant with the Hot-CAP is 29.1%, compared to 26.2% with the MEA process.

### **Cost Performance**

The capital and operation & maintenance (O&M) costs for the subcritical PC power plant equipped with the Hot-CAP were estimated by using the methodology introduced in ATTACHMENT 3 of the DOE/NETL DE-FOA-0000403. The financial model used for this study followed the same methodology as the NETL/DOE-2010/1397 report. The mass/energy calculation and equipment sizing results obtained from the process simulation were used for the cost estimation.

Cost results reveal that the Hot-CAP has a slightly lower capital cost (\$48 million) for CO<sub>2</sub> compression compared to benchmark MEA-based PCC (\$50 million), because the stripped product gas has a higher pressure. It also has a lower capital cost in the CO<sub>2</sub> capture section (\$351 million) compared to MEA (\$443 million), because it requires a smaller stripping column and does not employ an upstream polishing scrubber for deep sulfur removal. Overall, the total capital cost for the Hot-CAP, including CO<sub>2</sub> compression, is \$399 million, compared to \$493 million for the MEA process.

In addition, total O&M costs for the PC power plant with Hot-CAP is estimated to be \$176 million per year, slightly less than that of the PC plant with MEA (\$178 million/year). The loss of K<sub>2</sub>CO<sub>3</sub> solvent incurred by the reaction with SO<sub>2</sub> is counted in the cost analysis.

The 20-year levelized cost of electricity (LCOE) for the PC power plant with Hot-CAP, including CO<sub>2</sub> transportation and storage, is 119.6 mills/kWh, which is a 59% increase over the base PC plant without CO<sub>2</sub> capture. The LCOE increase caused by the Hot-CAP for CO<sub>2</sub> capture is 31% lower than the benchmark MEA-based process.

## 1 Introduction

The University of Illinois at Urbana-Champaign (UIUC) and Carbon Capture Scientific LLC (CCS LLC) have been developing an absorption-based technology for post-combustion CO<sub>2</sub> capture (referred to as Hot-CAP). The Hot-CAP technology employs a carbonate salt, such as potassium carbonate (K<sub>2</sub>CO<sub>3</sub>) or sodium carbonate (Na<sub>2</sub>CO<sub>3</sub>), as a solvent. The process involves four major unit operations - CO<sub>2</sub> absorption into a high-concentration carbonate solution, crystallization of bicarbonate, bicarbonate slurry-based high pressure CO<sub>2</sub> stripping, and reclamation of a sulfate salt from the sulfur dioxide (SO<sub>2</sub>) removal. The process has advantages of low energy use, minimal solvent degradation, and low solvent cost compared to conventional monoethanolamine (MEA)-based processes.

Under the DOE's Capture Program, a technology and economic feasibility study is required as a deliverable in the project Statement of Project Objectives (SOPO). This study analyzes a fully-integrated pulverized coal (PC) power plant equipped with the Hot-CAP technology for post-combustion CO<sub>2</sub> capture (PCC), and is carried out, to the maximum extent possible, in accordance to the methodology and data provided in ATTACHMENT 3 – Basis for Technology Feasibility Study of DOE Funding Opportunity Number: DE-FOA-0000403.<sup>[1]</sup>

The DOE/NETL report on “*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity* (Revision 2, November 2010), NETL Report No. DOE/NETL-2010/1397<sup>[2]</sup>” was used as the main source of reference to be followed, as per the guidelines of ATTACHMENT 3 of DE-FOA-0000403. CCS LLC adopted the design and economic evaluation basis from Case 10 of the DOE/NETL report. This case corresponds to a nominal 550 MWe (net) subcritical PC power plant that utilizes a benchmark 30 wt% MEA-based absorption system for post-combustion CO<sub>2</sub> capture.

For this techno-economic analysis (TEA), the Hot-CAP process replaces the MEA-based CO<sub>2</sub> absorption system in the original case. The objective of this study is to assess the performance of Hot-CAP fully-integrated with a subcritical PC plant similar to Case 10 of the DOE/NETL report, such that it corresponds to a nominal 550 MWe subcritical PC plant with 90% CO<sub>2</sub> capture. This plant has the same boiler firing rate and generated steam pressure as the DOE/NETL report's Case 10 PC plant. However, due to the difference in performance between the Hot-CAP and benchmark MEA-based CO<sub>2</sub> absorption technology, the net power output fluctuates.



- Wet bulb temperature, °C 10.8
- Ambient relative humidity, % 60

#### 2.1.4 Technical Assumptions and Data

Other technical data and assumptions include:

- Design coal feed to the power plant is Illinois No. 6. The coal properties are listed in Table 2.1 according to NETL’s Coal Quality Guidelines.

Table 2.1. Illinois No. 6 coal properties

Rank	Bituminous
Seam	Illinois #6 (Herrin)
Source	Old Ben Mine
Ultimate Analysis (as received), weight%	
Carbon	63.75
Hydrogen	4.5
Nitrogen	1.25
Chlorine	0.29
Sulfur	2.51
Oxygen	6.88
Ash	9.7
Moisture	11.12
Total	100
Proximate Analysis (as received), weight%	
Volatile Matter	34.99
Fixed Carbon	44.19
Ash	9.7
Moisture	11.12
Total	100
HHV (kJ/kg)	27,135

- Selected flows and operating conditions for the turbine are listed below:
  - Turbine gross power output, MW 673
  - SH HP steam inlet flow, 1000 kg/hr 2,364
  - HP turbine inlet pressure, MPa 16.65
  - HP turbine inlet temperature, °C 566
  - HP turbine outlet pressure, MPa 4.28
  - IP turbine inlet pressure, MPa 3.90
  - IP turbine inlet temperature, °C 566
  - IP turbine outlet pressure, MPa 0.51
  - LP turbine inlet pressure, MPa 0.51
  - Surface condenser pressure, mm Hg 50.8

- To generate the 2,364,000 kg/hr of SH HP steam to the STG, the boiler will burn 278,956 kg/hr of as-received Illinois No. 6 coal. The boiler firing rate and the SH HP steam generation rate will be held constant for the PCC case.
- Auxiliary loads for the overall plant can be separated into three categories: PCC-independent PC plant aux loads, PCC-dependent plant aux loads, and PCC loads. PCC-independent plant aux loads total 31,170 kWe with the breakdowns listed in Table 2.2.

PCC-dependent PC plant aux loads include cooling water (CW) circulation pump loads, cooling tower (CT) fan loads, and transformer loss. PC plant CW and CT loads are proportional to the STG surface condenser duty which varies with the PCC steam extraction requirement. Transformer loss is proportional to STG gross power output which also varies with PCC steam extraction requirement.

PCC loads will vary depending on the PCC design and include power consumed in the CO<sub>2</sub> capture and compression processes, plus any new CW and CT consumptions due to the PCC cooling loads.

Table 2.2. PCC-independent PC plant auxiliary loads breakdowns

Auxiliary loads breakdowns	kWe
Coal Handling and Conveying	540
Pulverizers	4,180
Sorbent Handling & Reagent Preparation	1,370
Ash Handling	800
Primary Air Fans	1,960
Forced Draft Fans	2,500
Induced Draft Fans	12,080
SCR	70
Baghouse	100
Wet FGD	4,470
Miscellaneous Balance of Power Plant	2,000
Steam Turbine Auxiliaries	400
Condensate Pumps	700
<b>Total</b>	<b>31,170</b>

- It is assumed that the subcritical PC plant utilizes a mechanical draft, evaporative cooling tower, and all process blowdown streams are treated and recycled to the cooling tower. The design ambient wet bulb temperature of 10.8°C is used to achieve a cooling water temperature of 15.6°C using an approach of 4.7°C. The PC cooling water range is assumed to be 11.1°C. The cooling tower makeup rate was determined using the following conditions:
  - Evaporative losses of 0.8% of the circulating water flow rate per 5.6°C of range;
  - Drift losses of 0.001% of the circulating water flow rate;
  - Blowdown losses are calculated as follows:
 
$$\text{Blowdown Losses} = \text{Evaporative Losses} / (\text{Cycles of Concentration} - 1)$$

where cycles of concentration is a measure of water quality, and a mid-range value of 4 is chosen for this study

- Raw water makeup was assumed to be provided 50% by a publicly owned treatment works and 50% from groundwater.

### 2.1.5 Environmental/Emissions Requirements

Design emissions requirements and limits for the subcritical power plant with PCC in this study are listed in Table 2.3.

Table 2.3. Air emissions targets

Controlled Pollutant	kg/MWh
SO <sub>2</sub>	0.008
NO <sub>x</sub>	0.339
Particulate Matter (Filterable)	0.063
Hg	5.53E-6

Emission component NO<sub>2</sub> and SO<sub>2</sub> can potentially be further removed from the flue gas through non-reversible reactions with the Hot-CAP solvent used. NO and Hg are assumed to pass through the PCC recovery unit and to be released to the atmosphere with the treated flue gas. PM is assumed to be removed from the flue gas through water and absorption solvent scrubbing.

## 2.2 PCC Design Criteria

### 2.2.1 General

The PCC plant is designed as an integral part of the subcritical PC power plant to capture up to 90% of the CO<sub>2</sub> in the flue gas. It is assumed that all of the fuel carbon is converted to CO<sub>2</sub> in the flue gas. CO<sub>2</sub> is also generated from limestone in the FGD system, and 90% of the total CO<sub>2</sub> exiting the FGD absorber is subsequently captured in the PCC.

The projected largest-single train size equipment will be used to maximize economy-of-scale. Vessels exceeding transportation size limits (as specified in the Project Transportation Size Limitation section of this document) will be field fabricated. The equipment is designed for a 30-year plant life.

Rotating equipment critical to the continuous plant operation is spared. When sparing is not feasible, alternate operation will be identified to maintain continuous power plant operation.

### 2.2.2 Flue Gas Feed Specification

The PC plant boiler will be burning 278,956 kg/hr of as-received Illinois No. 6 coal to generate 2,364,000 kg/hr of SH HP steam to the STG based on Case 10 subcritical PC plant in the DOE/NETL-2010/1397 report. Flue gas prior to the vent stack after it exits the wet FGD before

is the design feed for the PCC plant. The corresponding flue gas feed composition and flow rate is listed in Table 2.4.

Table 2.4. Flue gas composition and CO<sub>2</sub> capture process operating conditions

Parameter	Unit	Value
Flue gas inlet temperature	°C	58
Flue gas feed pressure	MPa	0.10
Flue gas flow rate	kg/hr	3,213,261
Flue gas composition		
N <sub>2</sub>	vol%	67.94
O <sub>2</sub>	vol%	2.38
CO <sub>2</sub>	vol%	13.50
Ar	vol%	0.81
H <sub>2</sub> O	vol%	15.37
Total		100

### 2.2.3 Design CO<sub>2</sub> Product Specifications

Recovered CO<sub>2</sub> is delivered at the battery limit with the following specifications:

- Inlet pressure, MPa 15.3
- Inlet temperature, °C 26
- CO<sub>2</sub> concentration (dehydrated), % >99.99
- N<sub>2</sub> + Ar concentration, ppmv < 1,000 (revised for PCC processes)
- O<sub>2</sub> concentration, ppmv < 100 (revised for PCC processes)
- H<sub>2</sub>O, ppmv < 50 (revised for molecule sieve drying)

### 2.2.4 Utility Commodity Specifications

- Intermediate Low Pressure Steam

Intermediate low pressure (ILP) steam for PCC stripper reboiling can be extracted from the power plant to meet the following PCC boundary limit conditions:

Minimum pressure	As Required
Temperature, °C	Saturation temperature + 10

The ILP steam is assumed to be de-superheated to 10°C above saturation temperature to allow positive control of de-superheater condensate injection. Degree of ILP steam superheat can be varied to meet minimum de-superheater design requirement.

- Return Condensate

Reboiler steam condensate will be pumped back to the power plant hot at:

Minimum pressure, MPa	1.2
Temperature, °C	75

- Cooling Tower Water

Cooling water from the plant cooling towers is available at the following conditions:

Maximum supply temperature, °C	16
--------------------------------	----

Maximum return temperature, °C	38
Maximum supply pressure, MPa	0.48
Maximum PCC pressure drop, MPa	0.21

### *2.2.5 Process Water Streams*

The PCC plant is designed to minimize/eliminate discharging hydrocarbon solvent-containing waste waters.

### 3 Simulations and Design of a Hot Carbonate Absorption Process with Crystallization-Enabled High Pressure Stripping

#### 3.1 Overview and Description of Hot-CAP Process

Figure 3.1 is a schematic diagram of the Hot-CAP. In this process, the flue gas from the baghouse or FGD of the power plant is directly introduced into the absorption column operating at 60-70°C and atmospheric pressure, where CO<sub>2</sub> and other acid gases are absorbed into a potassium carbonate (K<sub>2</sub>CO<sub>3</sub>) solution. The CO<sub>2</sub>-rich carbonate solution exiting the absorption column is cooled through a cross-flow heat exchanger by the CO<sub>2</sub>-lean carbonate solution returning from the crystallization tank. After passing the cross-flow heat exchanger, the CO<sub>2</sub>-rich carbonate solution enters the crystallization tank, where potassium bicarbonate (KHCO<sub>3</sub>) salt crystals are formed due to the low solubility of the bicarbonate at low temperatures (30-40°C). The crystals are separated and the resulting slurry is heated by the warmer regenerated lean carbonate solution from the stripper through another cross-flow heat exchanger prior to entering a high pressure stripper. The stripper operates at a high pressure (up to 10 atm) and high temperature (140-200°C). The CO<sub>2</sub> stream released in the stripper contains a relatively small amount of water vapor. The CO<sub>2</sub>-rich gas stream exiting the stripper is further cooled, dehydrated, and compressed to a sequestration-ready pressure. The CO<sub>2</sub>-lean solution exiting the bottom of the stripper enters the crystallization tank after exchanging heat with the feed slurry.

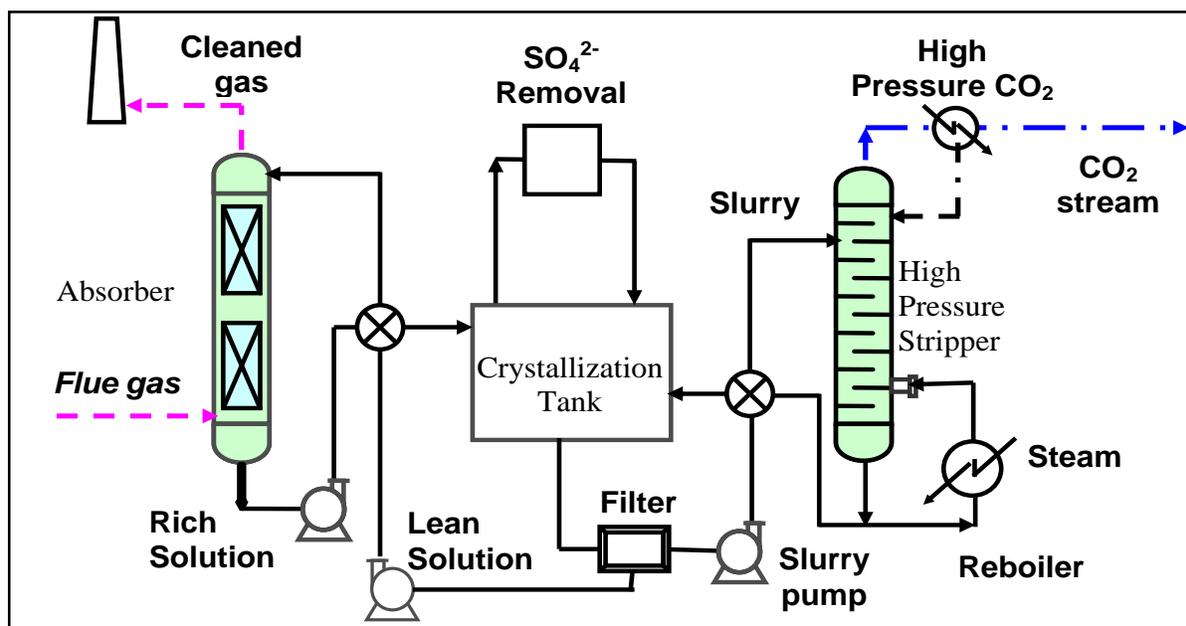


Figure 3.1. Schematic diagram of the proposed Hot-CAP.

The composition of the CO<sub>2</sub>-rich stream from the absorption column is 35~40 wt% (K<sub>2</sub>CO<sub>3</sub>-equivalent) solution with about 50% carbonate-to-bicarbonate (CTB) conversion. After the KHCO<sub>3</sub> crystallization, the CTB conversion level of the lean stream is 20% or less, which returns to the absorption column. The concentration of KHCO<sub>3</sub> in the absorption and crystallization process is subject to its solubility under different conditions, as shown in Figure 3.2.

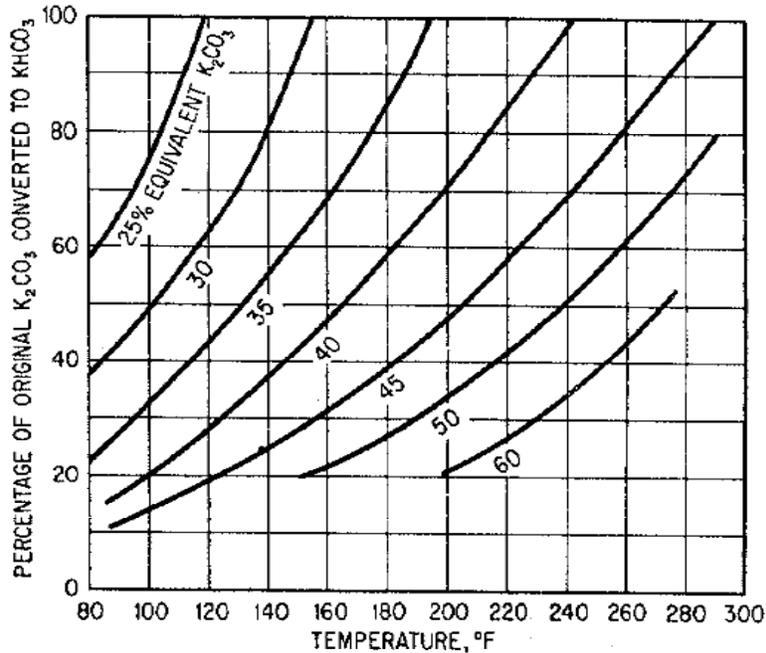


Figure 3.2. Effects of temperature and CTB conversion on the solubility of  $KHCO_3$  in  $K_2CO_3/KHCO_3$  solutions.

### 3.2 Risk Analysis and Mitigation Strategy for the Hot CAP Process

As part of the risk mitigation strategy analysis required by the DOE/NETL, CCS LLC performed a technology-focused risk analysis to identify critical technical risks and mitigate them through experiments, literature analysis, and discussion with equipment vendors. Five major technical risks were identified. The major technical risks are outlined in Figure 3.3.

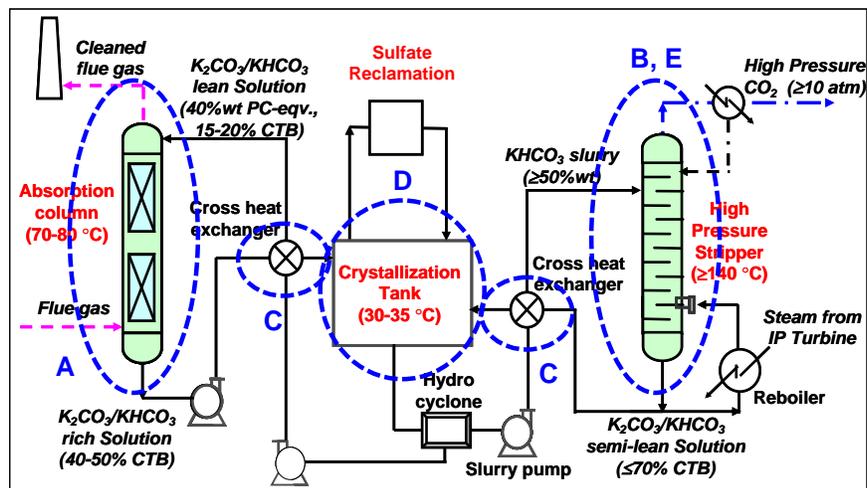


Figure 3.3. Five major technical risks identified for the Hot-CAP.

Risk A is related to the rate of  $CO_2$  absorption at elevated temperature ( $60-80^\circ C$ ) and concentration of carbonate solution ( $40-50$  wt%,  $K_2CO_3$ -equivalent). Risk B is related to the desired stripping pressure. The mitigation measures for these risks (A and B) were addressed

through the experimental and process simulation studies. Risks *C* and *D* are related to the design of the heat exchanger and crystallizer, and Risk *E* is related to the design of the high pressure stripping column and the related accessories. It was determined that risks (*C*, *D* and *E*) could be addressed through literature search, consultation from equipment vendors and design companies, and equipment design analysis. Details of these risks and methods to mitigate the risks are shown in Table 3.1.

Table 3.1. Technical risks of Hot-CAP and mitigation strategies

<b>Risk</b>	<b>Mitigation</b>	<b>Risk ID</b>
Rate of CO <sub>2</sub> absorption at temperature (60-80°C) and concentration of K <sub>2</sub> CO <sub>3</sub> solution (~40 wt%) insufficient to achieve process economics	Develop absorption promoters/catalysts and/or reconfigure absorption column design	A
Stripping pressure of potassium bicarbonate slurry is <10 atm, thereby unfavorably impacting process economics	Develop a sodium bicarbonate-based slurry in order to obtain stripping pressures ≥ 10 atm.	B
Heat exchanger fouled by slurry streams	Literature search, vender consultation, and engineering analysis to identify means to alleviate fouling	C
Crystallizer must be quickly cooled to achieve process economics	Literature search, vender consultation, and engineering analysis to identify means to achieve fast cooling in large systems	D
Commercially-available strippers require modifications to handle slurry and operate at high pressure	Literature search, vender consultation, and engineering analysis to determine means to modify standard stripper design	E

### 3.2.1 Risk A Mitigation Strategy

Experimental and simulation studies were performed to address Risk A. The experimental study was conducted at the UIUC to screen promoters that can accelerate the rate of absorption. The promoters were evaluated by the measured CO<sub>2</sub> removal efficiency by the promoted 40 wt% K<sub>2</sub>CO<sub>3</sub> solution in an absorption column. The selected experimental results are illustrated in Figure 3.4.

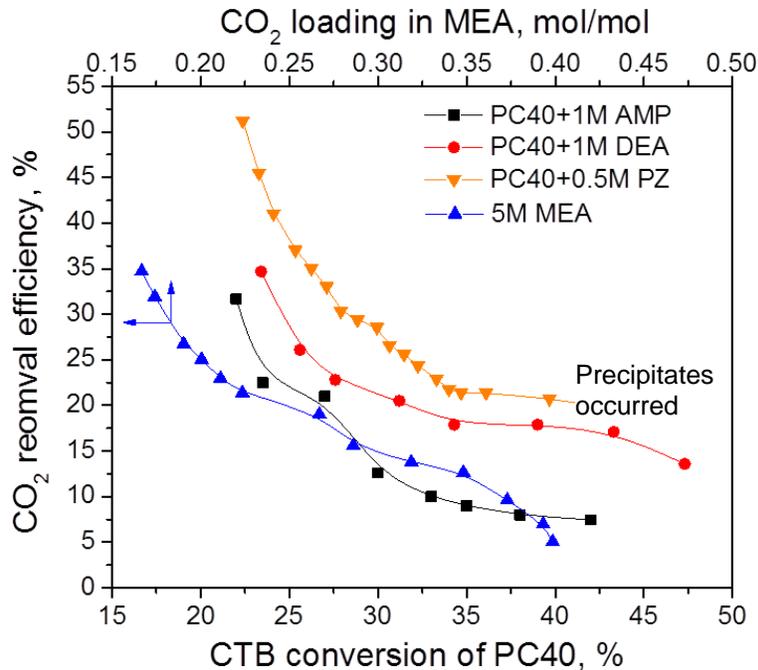


Figure 3.4. Experimental results of CO<sub>2</sub> absorption performance in the 40 wt% K<sub>2</sub>CO<sub>3</sub> solution with the addition of various promoters.

Details of the experimental study were described in a previous quarterly report submitted in January 2013. The key results from the experimental study include:

- CO<sub>2</sub> removal efficiency was low in the absence of a promoter; and
- CO<sub>2</sub> removal efficiency by the 40 wt% K<sub>2</sub>CO<sub>3</sub> solution promoted with 1M DEA or 0.5M PZ, either for the CO<sub>2</sub> lean or rich solution, was higher than that of the 5M MEA counterpart solution under the same operating conditions.

Simulations for the absorption process were performed in order to evaluate the performance of CO<sub>2</sub> absorption into carbonate solutions. Both thermodynamic and kinetic behaviors of the CO<sub>2</sub> absorption with or without a promoter were modeled. ChemCad software [3] was used for equilibrium-based process simulations, and ProTreat software [4] for rate-based simulations. Flue gas conditions were based on a 550 MWe subcritical PC plant referring to Case 10 of the DOE/NETL's Cost and Performance Baseline, [2] as shown in Table 2.4.

Simulation results of CO<sub>2</sub> absorption into K<sub>2</sub>CO<sub>3</sub> solutions at 70°C are summarized in Figures 3.5, 3.6, 3.7 and 3.8. The results demonstrate that the CO<sub>2</sub> removal efficiency was greatly increased by the addition of DEA or PZ promoter, which were consistent with the experimental findings. The following conclusions were drawn from the simulation study:

- The thermodynamic analysis indicates that the absorption of CO<sub>2</sub> into the 40 wt% K<sub>2</sub>CO<sub>3</sub> solution with a CO<sub>2</sub> loading equivalent to 15% initial CTB conversion is able to achieve the targeted 90% CO<sub>2</sub> removal at L/G ratios above 7. However, the absorption of CO<sub>2</sub> into the K<sub>2</sub>CO<sub>3</sub> with 20% initial CTB conversion cannot achieve the targeting 90% CO<sub>2</sub>

removal within a reasonable range of L/G ratios. In addition, a high  $K_2CO_3$  concentration is favorable for  $CO_2$  removal.

- The kinetic analysis of  $CO_2$  absorption into  $K_2CO_3$  solution without a promoter demonstrates that the  $CO_2$  removal efficiency is much less than the equilibrium value within a feasible range of column heights. The cost is high for absorbing  $CO_2$  using the  $K_2CO_3$  solution without a promoter.
- The kinetic analysis reveals that the  $CO_2$  removal efficiency can be greatly increased by the use of either DEA or PZ promoter. The  $CO_2$  removal efficiency increases with increasing promoter concentration. PZ promoter is more effective than DEA. In addition, 90%  $CO_2$  removal efficiency can be achieved using 40 wt%  $K_2CO_3$  with 20% initial CTB conversion promoted by 0.5M PZ at 60°C.

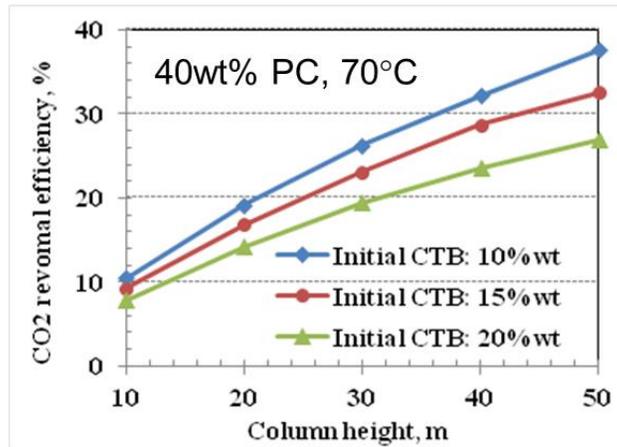


Figure 3.5 Simulation results of  $CO_2$  absorption into 40 wt%  $K_2CO_3$  solution without a promoter.

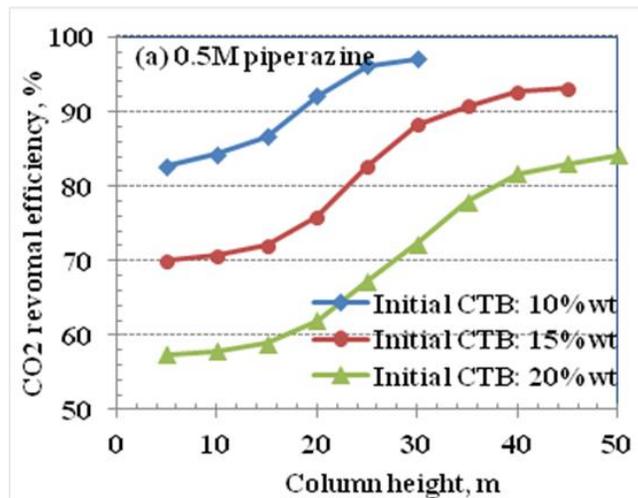


Figure 3.6. Simulation results of  $CO_2$  absorption into 40 wt%  $K_2CO_3$  with the addition of 0.5M PZ as a promoter.

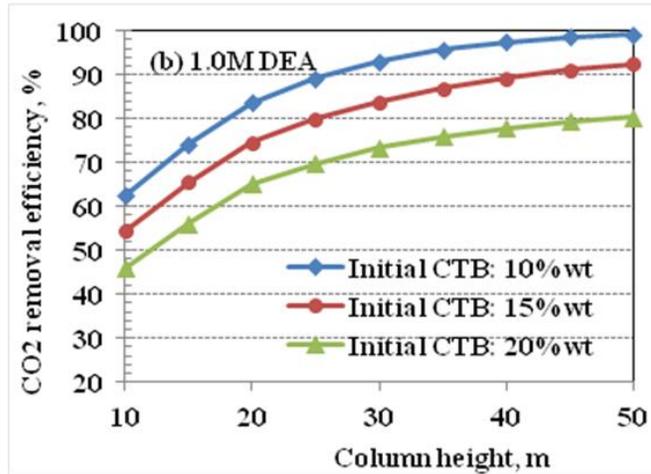


Figure 3.7. Simulation results of CO<sub>2</sub> absorption into 40 wt% K<sub>2</sub>CO<sub>3</sub> with the addition of 1.0M DEA as a promoter.

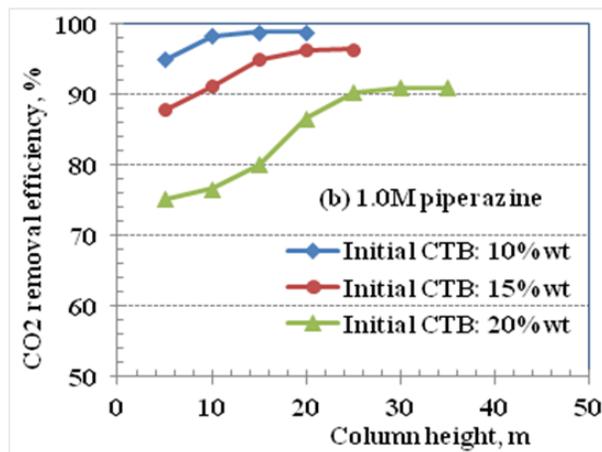


Figure 3.8. Simulation results of CO<sub>2</sub> absorption into 40 wt% K<sub>2</sub>CO<sub>3</sub> with the addition of 1.0M PZ as a promoter.

### 3.2.2 Risk B Mitigation Strategy

Risk *B* is addressed by the UIUC team based on the experimental results obtained from the phase equilibrium measurements and testing of CO<sub>2</sub> stripping in a high pressure stripping column. The stripper design is critical since a high stripping pressure and a smaller water vapor/CO<sub>2</sub> partial pressure ratio will significantly reduce the stripping heat (associated with water vaporization) during CO<sub>2</sub> stripping and the required compression work downstream. The measured VLE data confirmed the feasibility of high pressure CO<sub>2</sub> stripping in the Hot-CAP process. A higher stripping temperature, a higher level of CTB conversion, and a higher K<sub>2</sub>CO<sub>3</sub> concentration led to a higher stripping pressure and a lower water vapor/CO<sub>2</sub> ratio. However, recent studies at CCS LLC revealed that CO<sub>2</sub> stripping under excessively high pressure has the following drawbacks that must be considered:

- The required excessively high temperature steam reduces the net electric power generation; and
- Extra power is consumed for pumping the circulation solvent to higher pressures.

On the other hand, a low operating pressure results in a high stripping heat requirement. Therefore, the optimal stripping pressure is recommended to range between 5 and 10 bar.

### *3.2.3 Risk C Mitigation Strategy*

Risk *C* encompasses fouling caused by the need to manage slurry streams. Discussions with vendors indicated that fouling of the cross-flow heat exchangers and of the cooler inside the crystallizer due to possible potassium bicarbonate scaling on equipment surfaces can be solved. There are a variety of engineering solutions to reduce fouling, including:

- Reducing the temperature difference between the streams in the cross-flow heat exchangers;
- Pre-seeding of the crystallization solution;
- Using plate-and-frame type heat exchangers;
- Using a vacuum cooling crystallizer or a surface cooling crystallizer equipped with scrappers; and
- Adding extra heat exchangers

### *3.2.4 Risk D Mitigation Strategy*

In comparison, discussions with vendors related to Risk *D* indicated that the crystallizer design should be revised. Conventional crystallizer design requires a large temperature difference between the inlet solution (saturated or unsaturated carbonate solution entering the crystallizer) and the mother liquor (solution leaving the crystallizer). Therefore, the heat recovery from the incoming solution could be jeopardized if a single crystallizer configuration is used. Multiple Continuous Stirred Tank Reactor (CSTR) type crystallizers are required. A schematic of the revised design is shown in Figure 3.9. In this flowchart, five consecutive crystallization tanks are used instead of a single crystallizer (original design). The new configuration will reduce the temperature difference between the inlet and outlet streams in each crystallizer to about 5°C, thereby facilitating the heat recovery desired in the Hot-CAP process. This design was developed after numerous discussions with vendors.

### *3.2.5 Risk E Mitigation Strategy*

One of the major challenges in this project is the need to modify conventional strippers to handle slurry and operate at high pressure (Risk *E*). During the detailed analysis of Risk *E*, it was determined that there was an interaction between Risks *B* and *E* (i.e., high pressure stripping of the carbonate/bicarbonate slurry (Risk *B*), and the recrystallization of the bicarbonate during the cooling of the stripped lean solution (Risk *E*)). In the Hot-CAP process, bicarbonate needs to be regenerated at high pressure, which requires a combination of high total concentration of bicarbonate slurry and high CO<sub>2</sub> loading (high bicarbonate/carbonate ratio) for the regenerated lean solution. On the other hand, higher CO<sub>2</sub> loading in the stripped lean solution will bring recrystallization risk (Risk *E*) in the cooling process. This indicates a potential interaction between Risks *B* and *E*. In the subsequent process simulation and design, a reasonably high stripping pressure of 6 bar can be achieved using power plant steam at a relatively high temperature (about 180°C) as the heat source for the stripper reboiler.



Additionally, a reduced inlet temperature is beneficial for the PZ-promoted absorption reaction. Simulation results show that the addition of 1.0M PZ is required to promote the absorption at 70°C to achieve 90% CO<sub>2</sub> recovery, while 0.5M PZ is sufficient to achieve 90% CO<sub>2</sub> recovery at 60°C. On the other hand, a low inlet solvent temperature will significantly reduce the solubility of KHCO<sub>3</sub> in the rich solvent, which in turn, will reduce the solvent's working capacity. Therefore, the inlet temperature of the lean solvent is 60°C.

In summary, the hot carbonate solvent is a 40 wt% (K<sub>2</sub>CO<sub>3</sub>-equivalent) solution with 20% initial CTB conversion containing 0.5M PZ as a promoter. The design temperature of the inlet lean solvent is 60°C. Since absorption in the design solvent solution is sufficiently fast, the required flow rate of the solvent is limited by the KHCO<sub>3</sub> solubility in the rich solvent. Simulation results using ProTreat<sup>®</sup> show that the outlet temperature of the rich solvent reaches 67°C, at which, the solubility of KHCO<sub>3</sub> corresponds to 45% CTB conversion in the solution. At a working capacity equivalent to the CTB conversion varying from 20 to 45%, the required solvent flow rate is estimated to be 19,300 tonne/hr. Under these design conditions, the absorption column dimensions required for 90% CO<sub>2</sub> recovery are determined: the absorber consists of two parallel absorption columns, each with effective packing of 13-m in depth and 14.8-m in diameter.

### 3.3.2 Design of Stripping Column

It was found that both ChemCad and ProTreat<sup>®</sup> cannot provide sufficiently accurate performance predictions when the stripping temperature is higher than 140°C. However, the goal of this study is to achieve high pressure stripping (usually accompanying with a high temperature >140°C) to reduce the stripping heat loss and the required CO<sub>2</sub> compression work. Therefore, the stripping column simulation is based on a self-developed, steady-state thermodynamic model using the VLE data measured in this project. The experimental VLE data for CO<sub>2</sub> and water vapor are shown in Figures 3.10 and 3.11.

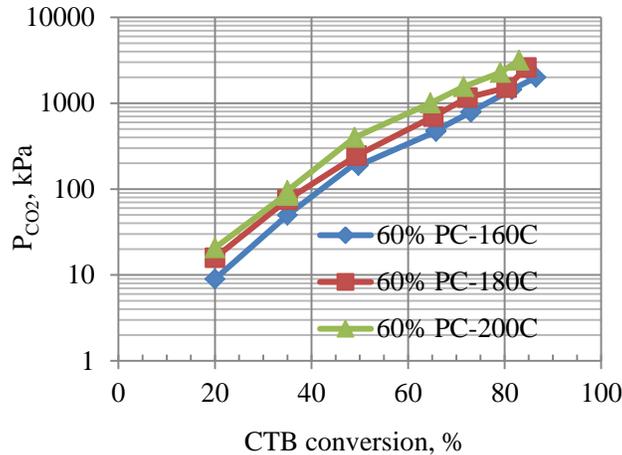


Figure 3.10. Experimental VLE data for CO<sub>2</sub> in 60 wt% K<sub>2</sub>CO<sub>3</sub> equivalent solution.

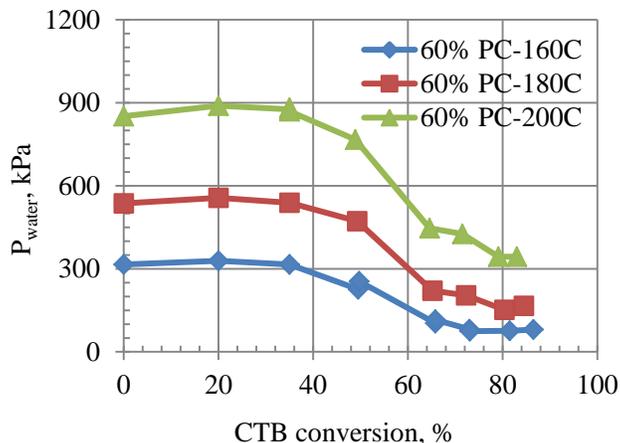


Figure 3.11. Experimental VLE data for water vapor in 60 wt%  $K_2CO_3$  equivalent solution.

The following design assumptions were applied to the simulation:

- Each stage is under ideal conditions and the vapor phase is in equilibrium with the liquid phase.
- The vapor phase consists of only  $CO_2$  and water vapor; any other components in the vapor phase are negligible.
- The rich solution entering the stripper is a 60 wt% ( $K_2CO_3$ -equivalent) solution, which is a blended slurry formed by potassium bicarbonate solids from crystallization tanks and a portion of the  $CO_2$ -rich solution from the absorption column, as shown in Figure 3.9. The CTB conversion in the rich solution is 79%.
- To reduce the potential risk of  $KHCO_3$  crystallization in the regenerated hot lean solution from the stripper when it is cooled during heat exchange (cross-flow heat exchanger #2 in Figure 3.9), the CTB conversion in the hot lean solution was kept at a level as low as possible.
- To reduce the stripping heat use, the temperature of the inlet rich solution was kept at a relatively low level. Based on the simulation using ProTreat<sup>®</sup>, the temperature difference between the top rich solution and the bottom lean solution is usually between 17-21°C.

Steady state simulation results by using the experimental VLE data indicate the CTB conversion in the hot lean solution can be reduced to 29% when the stripping pressure is maintained at 6 bar. The corresponding temperature in the stripper reboiler is 181°C. The corresponding temperature of the inlet rich solution at the top of the stripper is 161°C. From the difference in CTB conversion between the lean and rich solution, the flow rate of the 60 wt% rich solution can be estimated to be 7,094 tonne/hr.

The size of the stripping column was estimated using ProTreat<sup>®</sup>. As the VLE database built in ProTreat<sup>®</sup> is not available for temperatures above 140°C, the column sizing simulation was based on a 140°C stripping temperature at the bottom of the stripper with the reduced operating pressure of 2 bar. With the same flow rate and composition of the inlet solution obtained from the above steady state simulation, but with the inlet solution temperature reduced to 120°C, simulation results showed that the 29% CTB conversion in the hot regenerated lean solution can be achieved when the stripper is a single column of 10-m in height and 7.3-m in diameter. As the

CO<sub>2</sub> reaction kinetics usually increase with increasing temperature, the stripping column is conservatively sized using a single column with effective packing of 10-m in height and 7.3-m in diameter.

An important fact related to the stripper design is that all KHCO<sub>3</sub> solids in the inlet rich solution (79% CTB conversion and 161°C) entering the stripper is dissolved according to an exploration of the solubility data shown in Figure 3.2. Therefore, Risk E can be relieved from the stripper design.

### *3.3.3 Design of Crystallization Tanks*

Potassium bicarbonate crystallization is an important step in the Hot-CAP process. Risk analysis in Section 3.2.4 has revealed that a configuration of five consecutive stages of CSTR crystallizers, as shown in Figure 3.9, can be used instead of a single crystallizer to facilitate the heat recovery required in the process and reduce the temperature difference between the inlet and outlet streams of each crystallizer.

Based on an intensive literature review and discussions with vendors, it was concluded that a simple concrete tanker type of crystallizers with submerged coils can achieve the desired crystallization requirement. In addition, the cost of such crystallizers is the least expensive. Figure 3.12 illustrates a schematic of crystallizer structure, which has a draft-tube for internal circulation of magma and a downward-directed propeller agitator to provide a controllable circulation within the crystallizer. A part of the spiral heat exchange works as the draft-tube and the rest locates in the top region of the crystallizer. Both the top and bottom of the crystallizer are in conical shape so that the top region is able to provide a zone for fine crystal particles to settle. The clear mother solution leaves the crystallizer after overflowing to the next stage of crystallizer or return as a mother-liquor. Product slurry is removed through an outlet at the conical bottom and is further separated by a hydrocyclone. The separated liquid merges with the clear mother solution and leaves for the next crystallizer, or return as a mother liquor in the last crystallizer. The recovered crystal is used for preparing the inlet rich solution of the stripper.

The residence time of crystallizers and the size of crystal particles are the critical parameters for the design of crystallizers and crystal separators. The experimental results from the crystallization study conducted in this project show that KHCO<sub>3</sub> crystallization is instantaneous, but it takes about 30 minutes for the crystal particles to grow to 80 μm and above. In practical operation of a crystallizer, however, there is a large amount of crystal particles circulated in the crystallizer, which can be more than those recovered. Therefore, the residence time of the feed solution can be significantly reduced. In the current design, the average crystal solids concentration in each stage of the crystallizer is assumed at 10 wt%. The design volume of each crystallizer was estimated based on the crystallization kinetics obtained from the experimental results, as shown in Table 3.2.

On the other hand, a spiral tube heat exchanger soaked in a crystallizer to cool the inlet solution and remove the heat of KHCO<sub>3</sub> crystallization for each stage of crystallization, also occupies a part of the crystallizer volume. The required volume of the heat exchanger is assumed to be at least one third of the total volume of each crystallizer. There are two types of cooling medium used in a crystallizer. One is the returning lean solution, the mother liquor, for heat recovery, and

the other is external cooling water to remove the remaining cooling load to maintain the crystallizer operating at the desired temperature. The temperature approach to the crystallization temperature for the cooling lean solution is estimated using a logarithm mean temperature approach. A temperature change of cooling water from 17 to 32°C was adopted to determine the cooling water flow rate required for heat exchange in the crystallizer. As a result, the temperature approach for the external cooling water ranges from 17 to 38°C, depending on the temperature in the stage of crystallizer. The submerged coil heat exchanger is constructed with 5-cm diameter stainless steel tubing. The average heat transfer coefficient of the tube is 1,300 W/m<sup>2</sup>·K.

Table 3.2 Estimation of volumes for five stages of crystallizers

Stage of Crystallizer	1 <sup>st</sup>	2 <sup>nd</sup>	3 <sup>rd</sup>	4 <sup>th</sup>	5 <sup>th</sup>	Total
Feed flow, tonne/hr	22547	21592	20892	20356	19845	
Discharged crystal, tonne/hr	955	700	536	511	606	3308
Total cooling load, MW	88.6	143.9	127.7	123.6	159.3	643.1
Cooling by mother liquor, MW	74.5	73.4	101.7	0.0	0.0	249.6
Cooling by cooling water, MW	14.1	70.4	26.1	123.6	159.3	393.5
Crystallizer volume by cooling need, m <sup>3</sup>	309	355	293	155	269	1381
Design crystallizer volume, m <sup>3</sup>	360	540	640	760	920	3220
Actual residence time, sec	98	150	180	214	259	900
Cooling water flow, tonne/hr	757	3790	1404	6655	8577	21183

As shown in Figure 3.9, there are two sources of solutions required to precipitate out the KHCO<sub>3</sub> salt: 81% of the rich solution from the absorber (16,142 tonne/hr of 40 wt% K<sub>2</sub>CO<sub>3</sub> with 45% CTB conversion at 67°C), and all of the lean solution regenerated from the stripper (6,406 tonne/hr of 60 wt% K<sub>2</sub>CO<sub>3</sub> with 29% CTB conversion at 70°C). The density of the blended solution is estimated to be 1.71 kg/L at 68°C. The density of the returning mother liquor is 1.52 kg/L at 38°C.

The temperatures of the first to fifth crystallizer are controlled at 60, 55, 50, 45, and 38°C, respectively. The cooling load of each crystallizer is a sum of the solution's sensible heat and the KHCO<sub>3</sub> crystallization heat. The rate of KHCO<sub>3</sub> crystallization in each crystallizer was estimated based on its solubility, as shown in Figure 3.2. The total KHCO<sub>3</sub> crystallization heat released was estimated based on the crystallization rate and the molar crystallization heat (26.2 kJ/mole KHCO<sub>3</sub>). The sensible heat was estimated using ProTreat<sup>®</sup>. Table 3.2 summarizes the design parameters of the five crystallizers. It should be noted that some bicarbonate crystal particles are formed during heat exchange before the feed solutions enter the first crystallizer, especially for the lean solution from the stripping column that has a high carbonate equivalent concentration. This reduces the cooling load in the first crystallizer. The total volume of the five crystallizers is 3,220 m<sup>3</sup>, which results in an actual total residence time of 15 minutes. Thus, such a residence time is sufficient to meet both the heat exchange and crystal growth requirements. The total cooling water load is 393.5 MW, requiring a cooling water flow of 21,183 tonne/hr.

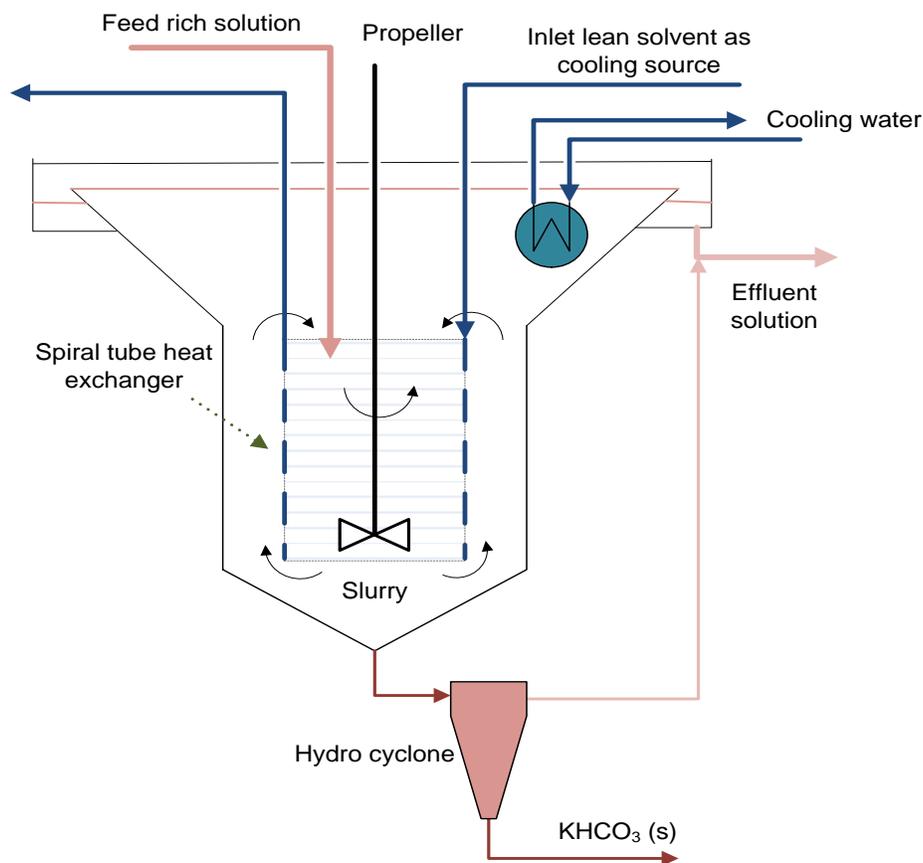


Figure 3.12. A schematic of crystallizer structure.

### 3.4 Mass and Heat Balance of Hot-CAP Process

Based on the flue gas composition and flow rate specified in Case 10 of the DOE/NETL-2010/1397 report, the mass and heat balances for the Hot-CAP are shown in Figure 3.13. The absolve is a 40 wt%  $K_2CO_3$  solution with 20% lean CTB conversion containing 0.5M PZ at a total flow rate of 19,300 tonne/hr. The  $CO_2$  loading of the solvent increases from 20% (lean) to about 45% (rich) CTB conversion. The rich  $K_2CO_3$  solvent from the absorber is separated into two streams - 81% of the rich solvent goes through the heat exchange and crystallization process, while the remaining part is mixed with  $KHCO_3$  crystal solids recovered from the crystallization tanks to produce a 60 wt% ( $K_2CO_3$ -equivalent) carbonate slurry with 79% CTB conversion.

In the crystallization process, 3,308 tonnes of  $KHCO_3$  solids are obtained from the blended solution made of 81% rich solvent from the absorber and all of the regenerated lean solvent from the stripper. The mother liquor from the crystallization process is mixed with a small amount of make-up water to form a 40 wt%  $K_2CO_3$  solution of 20% CTB conversion as the lean solution returning to the absorber. During crystallization, external cooling water is required to control the temperature of the different crystallizer units.

Although the crystallizers are designed to use covered tanks, water loss during the crystallization process is expected. The water loss rate is assumed at 0.1-0.2% of the solvent circulation rate.

The heat recovered from the first to the third crystallizers is sufficient to heat the lean solvent to the desired absorber inlet temperature (60°C). In total, 598 tonne/hr of CO<sub>2</sub> is released from the regeneration of KHCO<sub>3</sub> during the stripping process. After heat recovery from the regenerated hot lean solvent and the CO<sub>2</sub> product gas, 395.7 MW of external low pressure steam is required to meet the heat requirement, including 213.2 MW (thermal) in the reboiler and 182.5 MW (thermal) in the heat exchanger upstream of the stripping column. The total heat consumption in the Hot-CAP is estimated at 2,382 kJ/kg CO<sub>2</sub>.

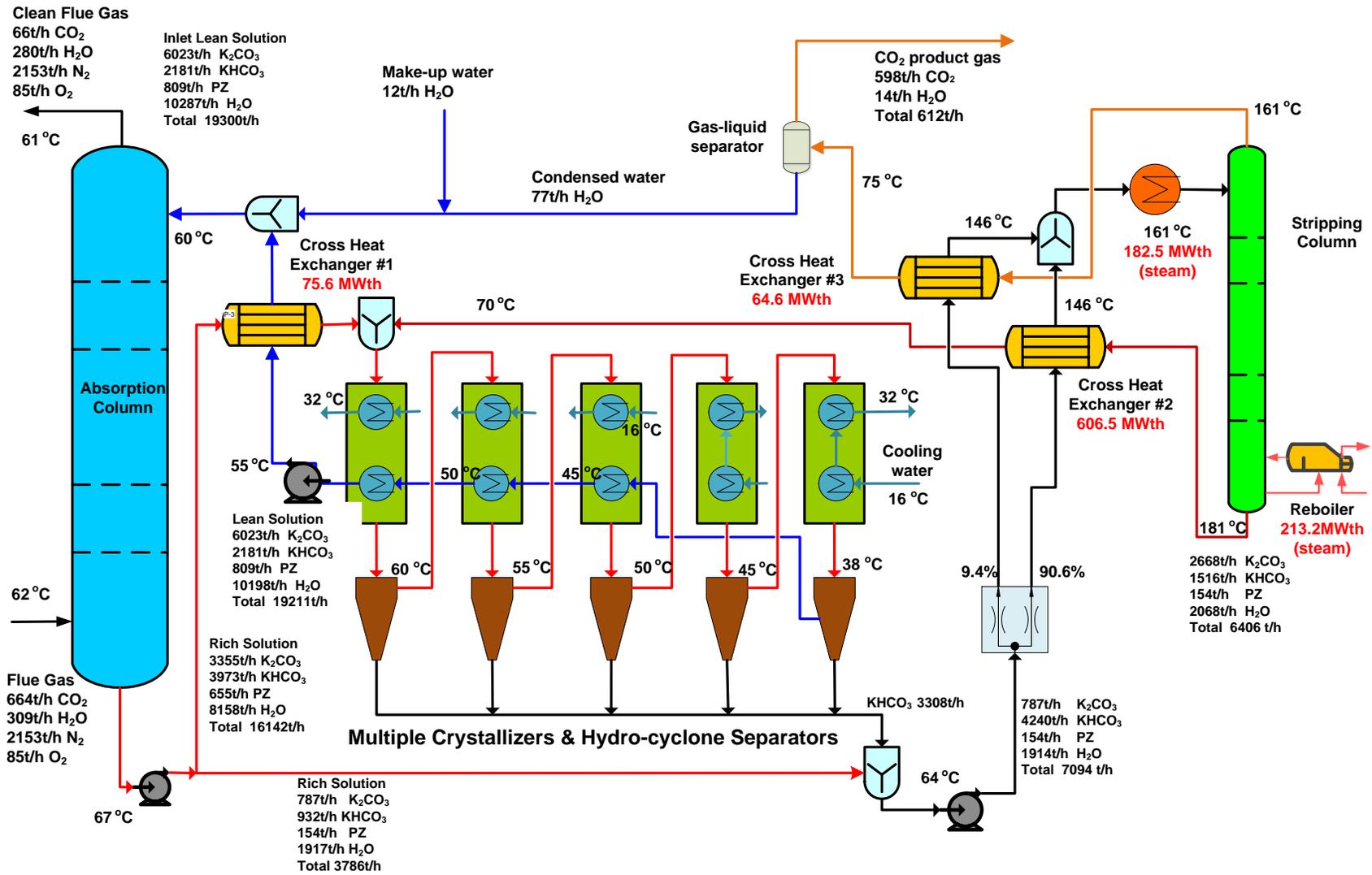


Figure 3.13. A schematic flowchart of the Hot-CAP process with mass and heat balances.

## 4 Cost Estimation

### 4.1 Cost Estimation Methodology

The Total Plant Cost (TPC) and Operation & Maintenance (O&M) costs for the subcritical PC power plant and the integrated CO<sub>2</sub> capture plant are estimated based on 2007 costs, using methodology introduced from ATTACHMENT 3 of the DOE/NETL DE-FOA-0000403<sup>[1]</sup>.

#### 4.1.1 Capital Cost

The DOE/NETL-2010/1397 report provided a cost estimate for 14 major subsystems of the Case 10 subcritical PC plant with CO<sub>2</sub> capture. Using this as the reference cost estimate, modifications to each subsystem were made either by capacity factoring, or by replacement with new estimates to obtain the overall cost estimate for the nominal 550 MWe subcritical PC plant with Hot-CAP based PCC. For the subsystems in which capacity factoring was used to perform the cost estimates, a power factor of 0.7 was applied. The list of the Case 10 subcritical PC plant subsystems and bases for modifications are shown in Table 4.1.

- Subcritical PC Plant

The capital cost estimates for the subcritical PC plant are developed based on the Case 10 costs provided in the DOE/NETL-2010/1397 report. The PCC section in this study differs from the CO<sub>2</sub> capture section provided in the report, resulting in a variation of the PC plant performance due to the differences in PCC design as well as solvent selection. As stated in Section 2.1.4, the revised PC plant with Hot-CAP-based PCC was estimated based on a different PCC LP steam extraction rate, hence resulting in a different power generation rate from the DOE/NETL Case 10 subcritical PC plant. For this reason, the PC plant equipment costs (primarily for the LP steam turbine, condenser and CW/CT sections) are re-estimated on a capacity-factor basis using the DOE/NETL reported costs as a baseline reference.

Material, direct labor, engineering and construction management fees, home office cost, and contingencies consistent with those used in the DOE/NETL report Case 10 are added to come up with the total subcritical PC plant cost estimate.

- PCC Plant

Capital cost for Hot-CAP-based PCC is a major equipment (ME) factored estimate for the DOE/NETL Case 10 subcritical plant with a target accuracy of  $\pm 30\%$ . Separate estimates are prepared for the CO<sub>2</sub> recovery facility and the CO<sub>2</sub> compression facility.

For an ME-factored estimate, ME material and labor costs were developed from equipment sizes, quantities, and design parameters defined by the PCC design discussed in the previous section. Bulk material and labor costs were factored from the ME costs. The sum of the ME and bulk material costs, including shipping costs, forms the total direct cost (TDC).

Construction indirect cost, factored from total direct labor cost, is added to the TDC to come up with the total field cost (TFC). Using factors consistent with the DOE/NETL report for the Case

10 TPC, the engineering and construction management fees, home office cost, and contingencies are added to the TFC to come up with the TPC.

Table 4.1. Cost estimate basis for subcritical PC plant with CO<sub>2</sub> capture

Acct. No.	Item/Description	Cost Estimate Basis	Capacity Factor Reference Basis
1	COAL & SORBENT HANDLING	Capacity Factor	AR Coal
2	COAL & SORBENT PREP & FEED	Capacity Factor	AR Coal
3	FEEDWATER & MISC. BOP SYSTEMS		
3.1	Feedwater System	Capacity Factor	AR Coal
3.2	Water Makeup & Pretreating	Capacity Factor	CW Makeup
3.3	Other Feedwater Subsystems	Capacity Factor	AR Coal
3.4	Service Water Systems	Capacity Factor	AR Coal
3.5	Other Boiler Plant Systems	Capacity Factor	AR Coal
3.6	FO Supply Sys & Nat Gas	Capacity Factor	AR Coal
3.7	Waste Treatment Equipment	Capacity Factor	AR Coal
3.8	Misc Equipment (Cranes, Air Comp, etc)	Capacity Factor	AR Coal
4	PC BOILER	Capacity Factor	AR Coal
5	FLUE GAS CLEANUP	Capacity Factor	AR Coal
5B	CO <sub>2</sub> REMOVAL & COMPRESSION		
5B.1	CO <sub>2</sub> Removal System	New estimate	N/A
5B.2	CO <sub>2</sub> Compression & Drying	New estimate	N/A
6	COMBUSTION TURBINE/ACCESSORIES	N/A	N/A
7	HRSG, DUCTING & STACK	Capacity Factor	AR Coal
8	STEAM TURBINE GENERATOR		
8.1	Steam TG & Accessories	Capacity Factor	STG Output
8.2	Turbine Plant Auxiliaries	Capacity Factor	STG Output
8.3	Condenser & Auxiliaries	Capacity Factor	Cond Duty
8.4	Steam Piping	Capacity Factor	Gross Power Output
8.9	TG Foundations	Capacity Factor	Gross Power Output
8.10	Back Pressure TG & Accessories	Capacity Factor	BPTG Output
9	COOLING WATER SYSTEM		
9.1	Cooling Towe	Capacity Factor	CT Load
9.2	Circulating CW Pump	Capacity Factor	CT Load
9.3	Circulating CW Syst Aux	Capacity Factor	CT Load
9.4	Circulating CW Piping	Capacity Factor	CT Load
9.5	Makeup Water System	Capacity Factor	CW Makeup
9.6	Closed CW System	Capacity Factor	CCW Load
9.9	Circ CW Syst Foundations & Structures	Capacity Factor	CT Load
10	ASH/SPENT SORBENT HANDLING SYS	Capacity Factor	AR Coal
11	ACCESSORY ELECTRIC PLANT	Capacity Factor	Gross Power Output
12	INSTRUMENTATION & CONTROL	Capacity Factor	AR Coal
13	IMPROVEMENT TO SITE	Capacity Factor	AR Coal
14	BUILDING & STRUCTURES	Capacity Factor	AR Coal

The absorption and stripping column dimensions were determined in the previous section. The dimensions of other major equipment used in the process are estimated based on the individual heat and material stream flows of the simulation.

Excluding the crystallization tanks, once the equipment was sized, the costs for the equipment and installation labor were estimated by scaling-up cost estimates for corresponding equipment in a TEA for "Post-Combustion Flue Gas CO<sub>2</sub> Capture Using Gas Pressurized Stripping Technology" performed by Nexant Inc for CCS LLC. The costs of crystallization tanks were estimated based on literature references and communication with vendors.

Costs for bulk materials, such as instrumentation, piping, structure steel, insulation, electrical, painting, and concrete and site preparation associated with the major equipment were factored from ME costs based on historical data for similar services. Installation labor for each bulk commodity was factored from historical data by type.

Construction indirect cost was factored from total direct labor costs based on historical data. Construction indirect cost covers the cost for setup, maintenance and removal of temporary facilities, warehousing, surveying and security services, maintenance of construction tools and equipment, consumables and utilities purchases, and field office payrolls.

Installation labor productivity and cost (wages, fringe benefit costs & payroll based taxes and insurance premiums) used to calculate the installation costs at 2007 price levels are based on estimation from the Nexant TEA.

- Engineering and Construction Management, Home Office Fees & Contingencies

Engineering and Construction Management are estimated as a percent of TFC. These costs consist of all home office engineering and procurement services, as well as field construction management costs.

Both the project contingency and process contingency costs represent costs that are expected to be spent in the development and execution of the project that are not yet fully reflected in the design. Project contingency is added to the TFC to cover project uncertainty and the cost of any additional equipment that would result during detailed design. Likewise, process contingency is added to the TFC to cover the cost of any additional equipment that would be required as a result of continued technology development. For this study, the factors used for the above fees and contingencies are consistent with those used in the DOE/NETL study.

#### *4.1.2 O&M Costs*

The O&M costs pertain to those charges associated with operating and maintaining the power plant over its expected life. These costs include:

- Operating labor
- Maintenance – material and labor
- Administrative and support labor
- Consumables
- Fuel
- Waste disposal

There are two components of O&M costs: fixed O&M, which is independent of power generation, and variable O&M, which is proportional to power generation. The variable O&M costs are estimated based on 85% capacity factor.

- Labor

Operating labor cost is determined based on the number of operators required to work in the plant. Other assumptions used in calculating the total labor cost include:

- 2007 Base hourly labor rate, \$/hr 33
- Length of work-week, hrs 40
- Labor burden, % 30
- Administrative/Support labor, % O&M Labor 25
- Maintenance material + labor, % TPC 1.64
- Maintenance labor only, % maintenance material + labor 40

- Consumables and Waste Disposal

The cost of consumables, including fuel, is determined based on the individual rates of consumption, the unit cost of each specific consumable commodity, and the plant annual operating hours. Waste quantities and disposal costs are evaluated similarly to the consumables. The unit costs for major consumables and waste disposal are based on the values reported in the DOE/NETL report. These costs are escalated to 2010, the year when construction is completed and production starts.

## 4.2 Financial Modeling Basis

The Financial Model used for this study follows the same methodology as the NETL/DOE-2010/1397 report. The method's figure-of-merit is the levelized cost of electricity (LCOE) over a 20-year period. The parameters to calculate the LCOE required by the model include the following:

- Income tax rate, % 38
- Percentage debt, % 45
- Interest rate, % 11
- Equity desired rate of return, % 12
- Repayment term of debt, years 15
- Depreciation 20 years, 150% declining balance
- Working capital None
- Plant economic life, years 30
- Tax holiday, years 0
- Start-Up costs (% of TPC less contingencies) 2
- EPC escalation, % per year 0
- Coal price nominal escalation, % 2.35
- O&M cost nominal escalation, % 1.87
- Duration of construction, years 3
- First year of construction 2007
- Construction cost distribution, %

- Year 1 5%
- Year 2 65%
- Year 3 30%

All costs are expressed in the “first-year-of-construction” year dollars, and the resulting LCOE is also expressed in “first-year-of-construction” year dollars. The DOE/NETL report’s net 550 MWe subcritical PC plant without CO<sub>2</sub> capture (Case 9) LCOE is to be used as the benchmark comparison for the subcritical PC plant with CO<sub>2</sub> capture. The Case 9 LCOE stated in the DOE/NETL report is 75.3 mills/kWh.

### **4.3 Performance Summary of Hot-CAP PCC**

Based on the mass and heat balances described in Section 3.4, an overall utilities sheet can be developed to summarize the Hot-CAP PCC process’ total steam requirement and electrical consumption. The process steam consumption is used to estimate the gross power generated by the power plant’s steam turbines.

The auxiliary loads for the overall plant are separated into three categories: PCC-independent PC plant auxiliary loads, PCC-dependent PC auxiliary loads, and PCC loads. The PCC-independent PC auxiliary loads are consistent with the values from the DOE/NETL report. The electrical load from the PCC utilities summary sheet is added directly to the total auxiliary loads as the PCC load. PCC-dependent PC aux loads, such as cooling water (CW) circulation pump loads, cooling tower (CT) fan loads, and transformer losses, vary with the PCC steam extraction requirement. These are calculated based on the PCC utilities consumption from the summary sheet and added to the total auxiliary load as the PCC-dependent PC auxiliary loads. Tables 4.2 and 4.3 show the utility consumption of the Hot-CAP-based PCC plant’s CO<sub>2</sub> capture section and CO<sub>2</sub> compression section, respectively.

Table 4.2. Hot-CAP-based PCC capture section utilities

Item No.	Item Name	Elec. Power	Steam		Water Cond.	Cooling Water		Regen. duty
			0.8MPa tonne/hr	1.3MPa tonne/hr		Load MW	Circ. rate tonne/hr	
<b>Exchangers</b>								
E-103	Stripping column inflow heater		247.3		(247.3)			182.5
E-104	Stripping column reboiler			274.8	(274.8)			213.2
E-105	Stage 1 crystallizer cooler					14.1	757	
E-106	Stage 2 crystallizer cooler					70.4	3,790	
E-107	Stage 3 crystallizer cooler					26.1	1,404	
E-108	Stage 4 crystallizer cooler					123.6	6,655	
E-109	Stage 5 crystallizer cooler					159.3	8,577	
<b>Pumps and drivers</b>								
G-101	Absorber bottom rich pump	709						
G-102	60 wt% slurry pump	2,338						
G-103	Absorber ovhd wash water pump	323						
G-104	Stage 1 crystallizer outlet pump	378						
G-105	Stage 2 crystallizer outlet pump	365						
G-106	Stage 3 crystallizer outlet pump	357						
G-107	Stage 4 crystallizer outlet pump	347						
G-108	Stage 5 crystallizer lean pump	2,795						
G-109	Stage 1 crystallizer agitator	462						
G-110	Stage 2 crystallizer agitator	1,272						
G-111	Stage 3 crystallizer agitator	1,945						
G-112	Stage 4 crystallizer agitator	2,989						
G-113	Stage 5 crystallizer agitator	4,819						
<b>Compressors</b>								
K-100	Flue gas blower	4,973						
Total		24,070	247.3	274.8	(522.1)	393.5	21,183	395.7

Table 4.3 Hot-CAP-based PCC CO<sub>2</sub> compression section utilities

Item No.	Item Name	Elec. Power	Steam	Water Condensate	Cooling Water		Regeneration duty
		kW	0.8MPa tonne/hr	Kg/hr	Load MW	Circ. rate tonne/hr	MW
<b>Exchangers</b>							
E-120	Stage 1 KO drum cooler				12.7	994	
E-121	Stage 2 KO drum cooler				11.7	916	
E-122	Stage 3 KO drum cooler				11.9	932	
E-123	Stage 4 KO drum cooler				13.9	1,088	
E-124	Stage 5 KO drum cooler				26.3	2,059	
<b>Compressors</b>							
K-101	Stage 1 CO <sub>2</sub> compressor	9,717					
K-102	Stage 2 CO <sub>2</sub> compressor	9,444					
K-103	Stage 3 CO <sub>2</sub> compressor	8,956					
K-104	Stage 4 CO <sub>2</sub> compressor	7,007					
K-105	Stage 4 CO <sub>2</sub> compressor	2,849					
<b>Packaged Equipment</b>							
V-100	TEG dehydration package	1,334	0.37	(0.37)	0.80	62	0.21
Total		39,307	0.37	(0.37)	77.3	6,052	0.21

#### 4.4 Capital Cost Estimate for Hot-CAP PCC

The Hot-CAP-based process' CO<sub>2</sub> capture and CO<sub>2</sub> compression section ME lists are provided in Tables 4.4 and 4.5, respectively. The estimated TFC for the CO<sub>2</sub> capture section (totaling \$226.2 million) and CO<sub>2</sub> compression section (totaling \$36.5 million), which includes the ME costs, freight, bulk materials and construction indirect costs, are shown in Tables 4.6 and 4.7, respectively. The TFC for the overall Hot-CAP PCC plant, totaling \$262.7 million, is listed in Table 4.8.

Table 4.4. Hot-CAP-based CO<sub>2</sub> capture section major equipment list

VESSELS & TANKS:													2007 Equip Cost \$1000			
Item No.	Item Name	Type	Design Conditions		Material of Construction	Quantity per Lot	Units	Inside Diameter m	Ht or Tan/Tan Length m	Width m	Length m	Number of Lots	Total Cost \$1000			
			kPa	°C												
C-101	Absorption Column	Vert	108	70	Kill CS	1	Vessel	14.8	24.0			2	25,253			
	- Sulzerpak	250.0 Sulzerf			304 SS	2236	cu.m					2				
	- Pall Rings	3.5 inch			304 SS	211	cu.m					2				
	- Support Plates				304 SS	344	sq. m	2 @ 14.8				2				
	- WW Support Plates				304 SS	106	sq. m	1 @ 11.6				2				
	- Hold-Down Plates				304 SS	344	sq. m	2 @ 14.8				2				
	- WW Hold-Down Plates				304 SS	106	sq. m	1 @ 11.6				2				
	- Liq Distributors				304 SS	344	sq. m	2 @ 14.8				2				
	- WW Liq Distributors				304 SS	106	sq. m	1 @ 11.6				2				
	- Chimney Trays				304 SS	344	sq. m	2 @ 14.8				2				
	- WW Chimney Trays				304 SS	106	sq. m	1 @ 11.6				2				
	- WW Demister Pads				304 SS	211	sq. m	2 @ 11.6				2				
C-102	Stripping Column	Vert	600	190	Kill CS	1	Vessel	7.3	20.0			1	8,273			
	- Sulzerpak	250.0 Sulzerf			304 SS	419	cu.m					1				
	- Support Plates				304 SS	84	sq. m	2 @ 7.3				1				
	- Hold-Down Plates				304 SS	84	sq. m	2 @ 7.3				1				
	- Liq Distributors				304 SS	84	sq. m	2 @ 7.3				1				
	- Chimney Trays				304 SS	84	sq. m	2 @ 7.3				1				
C-103	CO2 Product Gas Separator	Horizontal	600	75	304 SS	1	Vessel	2.3	2.6			1	119			
C-104	Stg 1 Crystallizer	Horizontal	101	60	Concrete	1	Tank	5.0	9.6			1	129			
C-105	Stg 2 Crystallizer	Horizontal	101	55	Concrete	1	Tank	5.0	11.7			1	193			
C-106	Stg 3 Crystallizer	Horizontal	101	50	Concrete	1	Tank	5.0	12.8			1	229			
C-107	Stg 4 Crystallizer	Horizontal	101	45	Concrete	1	Tank	5.0	13.9			1	271			
C-108	Stg 5 Crystallizer	Horizontal	101	38	Concrete	1	Tank	5.0	15.3			1	329			
C-109	Stage 1 crystallizer hydrocyclone	Cyclone	1000	60	CS	8	Vessel	0.3	1.0			1	80			
C-110	Stage 2 crystallizer hydrocyclone	Cyclone	1000	55	CS	6	Vessel	0.3	1.0			1	60			
C-111	Stage 3 crystallizer hydrocyclone	Cyclone	1000	50	CS	4	Vessel	0.3	1.0			1	40			
C-112	Stage 4 crystallizer hydrocyclone	Cyclone	1000	45	CS	4	Vessel	0.3	1.0			1	40			
C-113	Stage 5 crystallizer hydrocyclone	Cyclone	1000	38	CS	5	Vessel	0.3	1.0			1	50			
Sum												35,065				
SHELL & TUBE EXCHANGERS AND AIR COOLERS:													Total Equip Cost \$1000			
Item No.	Item Name	Type	Design P, kPa		Des Temp, °C		Material Of Construction		Duty MW	Total Bare Tube Area, sq. m		Physical Arrangement			Total # Req	Total Cost \$1000
			Shell	Tube	Shell	Tube	Shell	Tube		Series	In Parallel	In Parallel				
E-100	Absorber R/L HX #1	P&F	800	800	190	190	304SS	75.6	3086	1	3	3		3	370	
E-101	Stripper bottom R/L HX #2	P&F	800	800	190	190	304SS	606.5	13433	1	10	10		10	1,610	
E-102	CO2 product gas HX #3	P&F	800	800	190	190	304SS	64.6	1125	1	1	1		1	135	
E-103	Stripper inflow heater	P&F	800	800	190	190	304SS	182.5	3408	1	3	3		3	409	
E-104	Stripper reboiler	S&T	1300	800	190	190	304SS	213.2	13325	1	1	1		1	3,199	
E-105	Stage 1 crystallizer cooler by CW	STE		400		190	304SS	14.1	283	1	1	1		1	34	
E-106	Stage 2 crystallizer cooler by CW	STE		400		190	304SS	70.4	1632	1	2	2		2	196	
E-107	Stage 3 crystallizer cooler by CW	STE		400		190	304SS	26.1	713	1	1	1		1	85	
E-108	Stage 4 crystallizer cooler by CW	STE		400		190	304SS	123.6	4123	1	6	6		6	494	
E-109	Stage 5 crystallizer cooler by CW	STE		400		190	304SS	159.3	7165	1	10	10		10	859	
E-110	Stage 1 crystallizer cooler by Mother	STE		400		190	304SS	74.5	7962	1	10	10		10	954	
E-111	Stage 2 crystallizer cooler by Mother	STE		400		190	304SS	73.4	7846	1	10	10		10	941	
E-112	Stage 3 crystallizer cooler by Mother	STE		400		190	304SS	101.7	7109	1	10	10		10	852	
Sum												68	10,138			
COMPRESSORS, BLOWERS & DRIVERS:													Total Equip Cost \$1000			
Item No.	Item Name	Type	Design Conditions		Material Of Construction		Design Capacity			Pump BHP	Driver		Total # Req	Total Cost \$1000		
			kPa	°C	Wheel or Impel'r	Casing	Des Flow SCMH	Inlet kPa	Delta P kPa		HP	Type				
K-100	Flue Gas Blower	Cent.	200	60	304SS	304SS	2504320	102	6.0	3626	3817	Motor	2	2,539		
Sum												2	2,539			
PUMPS & DRIVERS:													Total Equip Cost \$1000			
Item No.	Item Name	Type	Design Conditions		Material Of Construction		Design Capacity			Pump BHP	Driver		Total # Req	Total Cost \$1000		
			kPa	°C	Wheel or Impel'r	Casing	Des Flow LPS	Inlet kPa	Delta P kPa		HP	Type				
G-101	Absorber bottom rich Pump	Cent.	300	70	CS	CS	3543	101	200	494	519	Motor	4	774		
G-102	Slurry pump	Cent.	1000	65	CS	CS	877	101	800	1178	1239	Motor	4	851		
G-103	Absorber Ovhd Wash Water Pump	Cent.	300	60	CS	CS	425	101	200	128	134	Motor	4	242		
G-104	Stage 1 crystallizer outlet pump	Cent.	300	60	CS	CS	3663	101	100	255	269	Motor	8	800		
G-105	Stage 2 crystallizer outlet pump	Cent.	300	55	CS	CS	3592	101	100	250	263	Motor	8	785		
G-106	Stage 3 crystallizer outlet pump	Cent.	300	50	CS	CS	3560	101	100	248	261	Motor	8	778		
G-107	Stage 4 crystallizer outlet pump	Cent.	300	45	CS	CS	3556	101	100	248	261	Motor	8	777		
G-108	Stage 5 crystallizer outlet pump	Cent.	300	38	CS	CS	3557	101	100	248	261	Motor	8	777		
G-109	Stage 1 crystallizer agitator	Propeller	1000	60	CS	CS	3663	400	10	619	650	Motor	1	148		
G-110	Stage 2 crystallizer agitator	Propeller	1000	55	CS	CS	3592	400	10	1705	1790	Motor	1	300		
G-111	Stage 3 crystallizer agitator	Propeller	1000	50	CS	CS	3560	400	10	2607	2738	Motor	1	404		
G-112	Stage 4 crystallizer agitator	Propeller	1000	45	CS	CS	3556	400	10	4007	4207	Motor	1	546		
G-113	Stage 5 crystallizer agitator	Propeller	1000	38	CS	CS	3557	400	10	6460	6783	Motor	1	763		
Sum												61	7945			
DUCTING													Total Equip Cost \$1000			
Item No.	Item Name	Type	Tube Design Cond		Insulation Thickness Inches	Duct Dimensions			Total # Req	Total Cost \$1000						
			kPa	°C		Mat Of Constr	Ht, m	Width, m			Total Length, m					
L-100	Flue Gas Feed & Exhaust Ducts	Duct	103	60	CS	1	4.5	4.5	1262	1	16731					
Sum										1	16731					
<b>TOTAL EQUIP COST</b>												<b>72417</b>				

Table 4.5. Hot-CAP-based compression section major equipment list

VESSELS & TANKS:													
=====													
Item No.	Item Name	Type	Design Conditions		Material of Construction	Quantity		Inside Diameter m	Ht or Tan/Tan Length m	Width m	Length m	Number of Lots	Total Equip Cost \$1000
			kPa	°C		per Lot	Units						
C-120	Stg 1 CO2 KO Drum	Vert	1,000	40	304Clad	1	Vessel	2.7	2.6			1	109
C-121	Stg 2 CO2 KO Drum	Vert	1,500	40	304Clad	1	Vessel	2.7	2.6			1	150
C-122	Stg 3 CO2 KO Drum	Vert	3,000	40	304Clad	1	Vessel	2.7	2.6			1	185
C-123	Stg 4 CO2 KO Drum	Vert	6,000	40	304Clad	1	Vessel	2.7	2.6			1	234
												Sum	678

SHELL & TUBE EXCHANGERS AND AIR COOLERS:														
=====														
Item No.	Item Name	Type	Design P, kPa		Des Temp, °C		Material Of Construction			Physical Arrangement			Total Equip Cost \$1000	
			Shell	Tube	Shell	Tube	Shell	Tube	Duty MW	Total Bare Tube Area, sq. m	In Series	In Parallel		Total # Req
E-120	Stg 1 KO Drum Cooler	S&T	1,000	600	190	190	304SS	304SS	12.7	1100	1	1	1	302
E-121	Stg 2 KO Drum Cooler	S&T	1,500	600	190	190	304SS	304SS	11.7	866	1	1	1	286
E-122	Stg 3 KO Drum Cooler	S&T	3,000	600	190	190	304SS	304SS	11.9	737	1	1	1	290
E-123	Stg 4 KO Drum Cooler	S&T	6,000	600	190	190	304SS	304SS	13.9	774	1	1	1	285
E-124	Stg 5 KO Drum Cooler	S&T	10,000	600	190	190	304SS	304SS	26.3	456	1	1	1	206
												Sum	1,369	

COMPRESSORS, BLOWERS & DRIVERS:														
=====														
Item No.	Item Name	Type	Design Conditions		Material Of Construction		Design Capacity			Driver			Total Equip Cost \$1000	
			kPa	°C	Wheel or Impel'r	Casing	Des Flow SCMh	Inlet kPa	Delta P kPa	Comp BHP	HP	Type		Total # Req
K-101	1st Stage CO2 Compressor	Cent.	2,000	40	CS	CS	308515	600	600.0	13025	13677	Motor	1	3,160
K-102	2nd Stage CO2 Compressor	Cent.	4,000	40	CS	CS	306589	1200	1200.0	12660	13292	Motor	1	2,120
K-103	3rd Stage CO2 Compressor	Cent.	6,000	40	CS	CS	305670	2400	2400.0	12005	12606	Motor	1	1,664
K-104	4th Stage CO2 Compressor	Cent.	9,500	40	CS	CS	305312	4800	4200.0	9393	9862	Motor	1	1,580
K-105	5th Stage CO2 Compressor	Cent.	16,600	40	CS	CS	305312	9000	6280.0	3819	4010	Motor	1	1,495
												Sum	10,020	

PUMPS & DRIVERS:													
=====													
Item No.	Item Name	Type	Design Conditions		Material Of Construction		Design Capacity			Driver			Total Equip Cost \$1000
			kPa	°C	Wheel or Impel'r	Casing	Des Flow LPS	Inlet kPa	Delta P kPa	Pump BHP	HP	Type	
												Sum	

PACKAGED & MISC EQUIPMENT:													
=====													
Item No.	Item Name	Type	Tube Design Cond		Mat Of Construct	Design Capacity	Remarks	Total # Req	Total Equip Cost \$1000				
			kPa	°C									
V-100	TEG Dehydration Package	Pkg						1	1722				
								Sum	1722				
								<b>TOTAL EQUIP COST</b>	<b>13788</b>				

Table 4.6. Hot-CAP-based PCC CO<sub>2</sub> capture section total field cost

COST CODE	DESCRIPTION	QTY	MEAS Unit	UNIT COSTS			D HIRE UNIT MH	TOTAL MHRS *		Equipment	COSTS IN U.S.\$1000			TOTAL
				MATL	LABOR	SC/Other		S/C	D HIRE		BULK	LABOR	SC/Other	
<b>PROCESS EQUIPMENT &amp; DUCTWORK</b>														
C	COLUMNS & TOWERS	3	EA							16,461		17,065		33,526
G	PUMPS & DRIVERS	61	EA							7,309		636		7,945
C	VESSELS, TANKS & STORAGE FACILITIES	6	EA							1,420		119		1,539
E	HEAT EXCHANGERS	68	EA							9,959		178		10,138
K	COMPRESSORS, BLOWERS, FANS & DRIVERS	2	EA							2,387		152		2,539
V	PACKAGED EQUIPMENT		EA											
L	DUCTWORK		EA								8,588	8,143		16,731
	FREIGHT	5.00	%							1,877				1,877
	TOTAL PROCESS EQUIPMENT & DUCTWORK		EA							39,412	8,588	26,294		74,294
	INSTRUMENTS													10,853
	PIPING													37,483
	STEELWORK													5,919
	INSULATION													4,605
	ELECTRICAL													19,399
	CONCRETE													7,343
	BUILDING													
	SITWORK													9,315
	PAINTING													460
	TOTAL OTHER DIRECT COSTS													95,377
	SUBTOTAL DIRECT COSTS													169,671
	SUBTOTAL CONSTRUCTION INDIRECT COSTS													56,500
	SUBTOTAL FIELD COSTS													226,171
	TOTAL (2007 BASIS)													226,171

Table 4.7. Hot-CAP-based PCC CO<sub>2</sub> compression section total field cost

COST CODE	DESCRIPTION	QTY	MEAS Unit	UNIT COSTS			D HIRE UNIT MH	TOTAL MHRS *		Equipment	COSTS IN U.S.\$1000			TOTAL
				MATL	LABOR	SC/Other		S/C	D HIRE		BULK	LABOR	SC/Other	
<b>PROCESS EQUIPMENT &amp; DUCTWORK</b>														
C	COLUMNS & TOWERS		EA											
G	PUMPS & DRIVERS	1	EA											
C	VESSELS, TANKS & STORAGE FACILITIES	4	EA							639		38		678
E	HEAT EXCHANGERS	5	EA							1,343		26		1,369
K	COMPRESSORS, BLOWERS, FANS & DRIVERS	4	EA							9,226		794		10,020
V	PACKAGED EQUIPMENT	1	EA							1,108		614		1,722
L	DUCTWORK		EA											
	FREIGHT	5.00	%							616				616
	TOTAL PROCESS EQUIPMENT & DUCTWORK		EA							12,933		1,472		14,405
	INSTRUMENTS													1,200
	PIPING													4,775
	STEELWORK													661
	INSULATION													647
	ELECTRICAL													3,976
	CONCRETE													1,959
	BUILDING													906
	SITWORK													1,357
	PAINTING													129
	TOTAL OTHER DIRECT COSTS													15,611
	SUBTOTAL DIRECT COSTS													30,016
	SUBTOTAL CONSTRUCTION INDIRECT COSTS													6,482
	SUBTOTAL FIELD COSTS													36,498
	<b>TOTAL (2007 BASIS)</b>													36,498



#### **4.5 Performance Summary of Subcritical PC Plant with Hot-CAP PCC**

According to the design of Hot-CAP-based PCC process described in Section 3, steam flows are required as heating source to meet the process needs. Given 10°C temperature approach, two steam flows with temperature 161°C and 181°C are designed to be extracted directly from the intermediate pressure (IP) turbine of the subcritical PC plant (Case 10 of the DOE/NETL report). Because the steam requirement for this Hot-CAP-based PCC plant is different from the benchmark MEA-based PCC in Case 10, there are inherent differences related to integration with the PC plant's steam cycle:

- The steam extraction rates for the PC plant with Hot-CAP-based PCC are 247 tonne/hr with 800 kPa pressure and 275 tonne/hr with 1,300 kPa pressure.
- The benchmark MEA-based PCC uses 876 tonne/hr with 507 kPa steam that is extracted at the Case 10 PC plant's IP/ LP crossover line.

Therefore, the net power output and thermal efficiency of the subcritical PC plant with Hot-CAP-based CO<sub>2</sub> capture differs from Case 10. Table 4.9 summarizes the performance and thermal efficiency of the overall PC plant with Hot-CAP-based PCC and provides a direct comparison to the benchmark MEA-based PCC. The net power output and efficiency of the subcritical PC plant with Hot-CAP CO<sub>2</sub> capture are 611 MWe and 29.1%, respectively, as compared to 550 MWe and 26.2% with MEA-based CO<sub>2</sub> capture.

Table 4.9. Performance summary of subcritical PC plant with Hot-CAP-based PCC

<b>Plant Performance Summary*</b>	<b>Case 10 with MEA-based PCC</b>	<b>Case 10 with Hot-CAP PCC</b>	
<b>Consumables:</b>			
As-Received Coal Feed, kg/hr	278,956	278,956	
Limestone Sorbent Feed, kg/hr	28,404	28,404	
Thermal Input, kWt	2,102,643	2,102,643	
<b>PCC Steam Extraction:</b>		Steam flow 1	Steam flow 2
Extraction Pressure, kPa	507	800	1300
Extraction Temperature, °C	296	316	388
Flow before DeSuperHt, 1000 kg/hr	876	247	275
Reboiler Condensate Return Temp, °C	348	170	191
<b>Power Generation Summary, kW:</b>			
Steam Turbine Gross Power	672,700	722,695	
<b>Auxiliary Load Summary, kW:</b>			
Coal Handling and Conveying	540	540	
Limestone Handling & Reagent Preparation	1,370	1,370	
Pulverizers	4,180	4,180	
Ash Handling	800	800	
Primary Air Fans	1,960	1,960	
Forced Draft Fans	2,500	2,500	
Induced Draft Fans	12,080	12,080	
SCR	70	70	
Baghouse	100	100	
FGD Pumps and Agitators	4,470	4,470	
Misc Balance of Plant	2,000	2,000	
Steam Turbine Auxiliaries	400	400	
Condensate Pumps	700	700	
Ground Water Pump	1,020	1,020	
Cooling Water Circulation Pumps	11,190	8,693	
Cooling Tower Fans	5,820	4,521	
Transformer Losses	2,350	2,600	
CO <sub>2</sub> Capture Plant Auxiliaries	22,400	24,070	
CO <sub>2</sub> Compression	48,790	39,307	
<b>Total Auxiliaries, kW</b>	<b>122,740</b>	<b>111,381</b>	
<b>Net Power Export, kW</b>	<b>549,960</b>	<b>611,314</b>	
<b>Net Plant Efficiency, % HHV</b>	<b>26.2</b>	<b>29.1</b>	
<b>Net Plant Heat Rate, kJ/kWh</b>	<b>13,764</b>	<b>12,404</b>	
<b>Cooling Tower Loads, GJ/hr:</b>			
Surface Condenser Duty	2,034	2,670	
CO <sub>2</sub> Capture Plant Cooling Duties	3,585	1,417	
CO <sub>2</sub> Compression Cooling Duties	Incl Above	278	
<b>Total Cooling Tower Loads, GJ/hr</b>	<b>5,619</b>	<b>4,365</b>	
<b>Overall Makeup Water Balance, m<sup>3</sup>/min:</b>			
FGD Makeup	5.5	5.5	
BFW Makeup	0.4	0.4	
CO <sub>2</sub> Capture & Compression Makeups	0.2	0.2	
Cooling Tower Makeup	36.4	28.3	
<b>Net Raw Water Makeup, m<sup>3</sup>/min</b>	<b>42.5</b>	<b>34.4</b>	

#### **4.6 Capital Cost Estimate for Subcritical PC Plant with Hot-CAP PCC**

Based on the cost estimating methodology for the overall subcritical PC plant with Hot-CAP-based PCC, as described in Section 4.1, Table 4.10 shows the total plant capital cost organized by cost account following the format of the DOE/NETL report. The engineering, construction management, and home office fees, as well as project and process contingencies are applied to the TFC (Bare Erected Cost in the DOE/NETL report) to arrive at the total subcritical PC plant with PCC capital cost. The total plant cost for the 550 MWe PC plant equipped with Hot-CAP for 90% CO<sub>2</sub> capture and compression is \$1,523 million.

Table 4.10. Capital cost estimate for the subcritical PC plant with Hot-CAP-based PCC

Hot CAP PCC Total Plant Cost Details in U.S.\$1000													
Acct No.	Plant Description	Capacity Scaling	Capacity	Equip Cost	Material Cost	Direct Labor Cost	Indirect Labor Cost	Sales Tax	Bare Erect Cost	Eng'g CM H.O. & Fee	Contingency		Total Pft Cost
		Units									Process	Project	
1	COAL & SORBENT HANDLING	= NETL Case 10		21,191	5,688	12,662	0	0	39,541	3,548	0	6,463	49,552
2	COAL & SORBENT PREP & FEED	= NETL Case 10		14,465	844	3,675	0	0	18,984	1,664	0	3,097	23,745
3	FEEDWATER & MISC BOP SYSTEMS												
	Feedwater System	= NETL Case 10		20,624	0	7,119	0	0	27,743	2,430	0	4,526	34,699
	Water Makeup & Pretreating	CW makeup, m <sup>3</sup> /min		7,889	0	2,539	0	0	10,428	991	0	2,284	13,703
	Other Feedwater Subsystems	= NETL Case 10		6,747	0	2,851	0	0	9,598	860	0	1,569	12,027
	Service Water Systems	= NETL Case 10		1,471	0	800	0	0	2,271	214	0	497	2,982
	Other Boiler Plant Systems	= NETL Case 10		8,081	0	7,979	0	0	16,060	1,526	0	2,638	20,224
	FO Supply & Nat Gas	= NETL Case 10		278	0	348	0	0	626	59	0	103	788
	Waste Treatment Equipment	= NETL Case 10		5,087	0	2,900	0	0	7,987	777	0	1,753	10,517
	Misc Equip (Cranes, Air Comp, etc)	= NETL Case 10		2,955	0	903	0	0	3,858	371	0	846	5,075
4	PC BOILER	= NETL Case 10		171,007	0	109,973	0	0	280,980	27,374	0	30,835	339,189
5	FLUE GAS CLEANUP	= NETL Case 10		107,581	0	36,768	0	0	144,349	13,816	0	15,817	173,982
5B1	Hot CAP CO2 REMOVAL SYSTEM	Independently Calc		39,412	73,770	112,989	0	0	226,171	21,486	45,234	58,578	351,470
5B2	CO2 COMPRESSION & DRYING	Independently Calc		12,933	10,793	12,772	0	0	36,498	3,467	0	7,993	47,958
6	COMBUSTION TURBINE/ACCESSORIES	kW GT output	N/A	N/A	N/A	N/A	0	0	0	0	0	0	0
7	HRSRG, DUCTING & STACK	= NETL Case 10		19,509	1,069	13,214	0	0	33,792	3,095	0	4,848	41,735
8	STEAM TURBINE GENERATOR												
	Steam TG & Accessories	= NETL Case 10		55,209	0	6,905	0	0	62,114	5,951	0	6,806	74,871
	Turbine Plant Auxiliaries	= NETL Case 10		387	0	828	0	0	1,215	119	0	133	1,467
	Condenser & Auxiliaries	MMBtu/h Cond Q		6,471	0	2,680	0	0	9,151	869	0	1,002	11,022
	Steam Piping	kW Total TG output		21,345	0	10,524	0	0	31,869	2,659	0	5,179	39,707
	TG Foundations	kW Total TG output		0	1,213	1,917	0	0	3,130	296	0	685	4,111
9	COOLING WATER SYSTEM												
	Cooling Tower	MMBtu/h CT Load	4,365	13,961	0	4,347	0	0	18,309	1,739	0	2,005	22,053
	Circulating CW Pump	MMBtu/h CT Load	4,365	2,905	0	218	0	0	3,123	264	0	339	3,726
	Circulating CW Syst Aux	MMBtu/h CT Load	4,365	703	0	94	0	0	797	76	0	87	960
	Circulating CW Piping	MMBtu/h CT Load	4,365	0	5,575	5,403	0	0	10,978	1,043	0	1,803	13,824
	Makeup Water System	CW makeup, m <sup>3</sup> /min	34	586	0	784	0	0	1,370	130	0	225	1,725
	Closed CW System	MMBtu/h CT Load	4,365	823	0	655	0	0	1,478	140	0	243	1,861
	Circ CW Syst Foundations & Structures	MMBtu/h CT Load	4,365	0	3,307	5,253	0	0	8,560	813	0	1,875	11,248
10	ASH/SPENT SORBENT HANDLING SYS	= NETL Case 10		5,525	176	7,387	0	0	13,088	1,258	0	1,477	15,823
11	ACCESSORY ELECTRIC PLANT	kW Gross output		27,283	11,626	32,922	0	0	71,832	6,393	0	9,966	88,190
12	INSTRUMENTATION & CONTROL	= NETL Case 10		9,942	0	10,082	0	0	20,024	1,816	1,001	2,805	25,646
13	IMPROVEMENT TO SITE	= NETL Case 10		3,344	1,922	6,739	0	0	12,005	1,184	0	2,638	15,827
14	BUILDING & STRUCTURES	= NETL Case 10		0	25,775	24,432	0	0	50,207	4,529	0	8,210	62,946
<b>TOTAL COST</b>				<b>587,715</b>	<b>141,757</b>	<b>448,663</b>			<b>1,178,135</b>	<b>110,959</b>	<b>46,235</b>	<b>187,325</b>	<b>1,522,653</b>

#### **4.7 O&M Cost Estimate for Subcritical PC Plant with Hot-CAP PCC**

According to Section 4.1, the annual O&M cost consists of two components: fixed O&M, which is independent of power generation, and variable O&M, which is proportional to power generation and is estimated based on 85% annual capacity factor.

The costs of consumables are escalated to 2010, the year when construction is completed. The annual escalation factor for all consumables, excluding fuel, is 1.87%. For the Illinois No 6 coal, the annual escalation factor is 2.35%.

The annual variable O&M costs, including consumables such as fuel, water and chemicals, as well as waste disposal costs, are determined based on the rates of consumption, the unit cost of each commodity, and total annual operating hours. Since half of the net raw water makeup comes from groundwater, water expenditure was calculated based on one-half of the net raw water makeup from the overall plant water balance. PC plant related chemical expenditures are independent of the PCC and follow the NETL/DOE report Case 10 consumptions.

PCC-dependent chemicals include: piperazine promoter, active carbon for piperazine filter, and  $K_2CO_3$  solvent. Different from Case 10, no sulfur polishing facility is used before  $CO_2$  capture in the Hot-CAP. Instead, the remaining  $SO_2$  in the flue gas from the FGD reacts directly with  $K_2CO_3$  in the absorber, resulting in degradation loss of  $K_2CO_3$  solvent. The  $K_2CO_3$  solvent replacement cost is thus estimated based on the total degradation rate caused by the reaction with  $SO_2$ , assuming a flue gas  $SO_2$  concentration of 42 ppmv entering the Hot-CAP absorber. A corrosion inhibitor is not required due to the assumption that Hot-CAP is less corrosive than MEA.

The total estimated annual O&M cost for the subcritical PC plant with Hot-CAP-based PCC is \$176 million. Table 4.11 shows the breakdown of the O&M costs in a similar format to the DOE/NETL Case 10 report.

Table 4.11. O&M costs of the subcritical PC plant with Hot-CAP-based PCC

<b>Fixed Operating Expenses:</b>				<b>Expense</b>
Annual Operating Labor Cost				\$6,134,700
Maintenance Labor Cost				\$9,988,606
Administration & Support Labor				\$4,030,826
Property Taxes and Insurance				\$30,453,066
<b>Total Fixed Operating Costs</b>				<b>\$50,607,198</b>
<b>Variable Operating Expenses:</b>				
Maintenance Material Cost				\$14,982,908
	<u>Consumables</u>	<u>Consumption/day</u>	<u>Unit Cost</u>	
Water(/1000 gallons)		6539	1.08	\$2,191,095
Chemicals:				
MU & WT Chemicals (lb)		39119	0.17	\$2,100,447
Limestone (ton)		751	21.63	\$5,043,346
Carbon for Mercury Removal (lb)		0	1.05	\$0
Piperazine Promotor (lb)		500	4.54	\$704,392
K <sub>2</sub> CO <sub>3</sub> Solvent (ton)		15.60	544.20	\$2,633,874
NaOH (ton)		7.89	433.68	\$1,061,704
H <sub>2</sub> SO <sub>4</sub> (ton)		7.53	138.78	\$324,217
Corrosion Inhibitor		0	0	\$0
Act Carbon (lb)		168	1.05	\$54,728
Ammonia (19% NH <sub>3</sub> )(ton)		110	129.8	\$4,446,378
SCR Catalysts (m3)		0.46	5775.94	\$831,516
Waste Disposal:				
Flyash (ton)		572	16.23	\$2,881,846
Bottom Ash (ton)		143	16.23	\$720,462
Gypsum		1,159	0	\$0
As-Received Coal Feed		7,380	38.18	\$87,425,787
<b>Total Variable Operating Costs</b>				<b>\$125,402,700</b>
<b>TOTAL ANNUAL OPERATING COSTS</b>				<b>\$176,009,897</b>

#### 4.8 LCOE Estimate for Subcritical PC Plant with Hot-CAP PCC

Using the methodology as described in Section 4.1, the estimated LCOE for the subcritical PC plant with Hot-CAP-based PCC and comparison to the corresponding values of DOE/NETL report Case 9 and 10 are listed in Table 4.12. The LCOE of the Hot-CAP PCC-based PC plant is 119.6 mills/kWh, which is 159% of the LCOE (75.3 mills/kWh) of the same PC plant without CO<sub>2</sub> capture. Compared to a 139.0 mills/kWh LCOE for the MEA PCC-based PC plant, the

increase of LCOE caused by CO<sub>2</sub> capture with the Hot-CAP is 31% lower than that with benchmark 30 wt% MEA.

Table 4.12. LCOE estimate for the subcritical PC plant with Hot-CAP-based PCC

<b>Post-Combustion Case Description</b>	Subcritical PC		
	w/o CO <sub>2</sub> Capture	w/ CO <sub>2</sub> Capture	
<b>Case Number</b>	<b>Case 9</b>	<b>Case 10</b>	<b>This Study</b>
<b>Type of CO<sub>2</sub> Capture Technology</b>	<b>N/A</b>	<b>Econoamine</b>	<b>Hot-CAP</b>
<b>Capital Cost Year</b>	2007	2007	2007
<b>CO<sub>2</sub> Capture</b>	0%	90%	90%
<b>Power Production, MW</b>			
Gross Power	583	673	723
Net Power	550	550	611
<b>Cost</b>			
<b>Total Plant Cost, 2007\$/kW</b>	1,662	2,942	2,491
<b>Total Overnight Cost, 2007\$/kW</b>	1,996	3,610	3,051
Bare Erected Cost	1,317	2,255	1,927
Home Office Expenses	124	213	182
Project Contingency	182	369	306
Process Contingency	0	105	76
Owner's Costs	374	667	560
<b>Total Overnight Cost, 2007\$ ×1000</b>	1,098,124	1,985,432	1,865,250
<b>Total As Spent Capital, 2007\$/kW</b>	2,264	4,115	3,478
<b>COE (mills/kWh, 2007\$)</b>	59.4	109.6	94.3
CO <sub>2</sub> TS&M Costs	0.0	5.8	5.2
Fuel Costs	15.2	21.3	19.2
Variable costs	5.1	9.2	8.3
Fixed Costs	7.8	13.1	11.1
Capital Costs	31.2	60.2	50.4
<b>LCOE, mills/kWh</b>	<b>75.3</b>	<b>139.0</b>	<b>119.6</b>
<b>% of Case 9 LCOE</b>	<b>100%</b>	<b>185%</b>	<b>159%</b>

## 5 Summary

A TEA was performed to compare the energy use and cost performance of a nominal 550 MWe subcritical PC plant without CO<sub>2</sub> capture (DOE/NETL Case 9) with benchmark MEA-based PCC (DOE/NETL Case 10) and Hot-CAP-based PCC. The results show that the net power produced in the PC plant equipped with Hot-CAP is 611 MWe, greater than that with MEA (550 MWe) due to a reduced steam demand for CO<sub>2</sub> stripping.

The Hot-CAP has a slightly lower capital cost (\$48 million) for CO<sub>2</sub> compression compared to MEA-based PCC (\$50 million), because the stripped product gas has a higher pressure (6 bar vs. 1.6 bar). The Hot-CAP also has a lower capital cost in the CO<sub>2</sub> capture section (\$351 million) compared to MEA-based PCC (\$443 million), because it requires a smaller stripping column and does not employ an upstream polishing unit for deep sulfur removal prior to CO<sub>2</sub> capture.

O&M costs for the PC plant equipped with Hot-CAP are estimated to be \$176 million annually, less than that of the PC plant with MEA (\$178 million). The loss of K<sub>2</sub>CO<sub>3</sub> solvent incurred by the reaction with SO<sub>2</sub> is included in the analysis.

The 20-year LCOE for the PC plant with Hot-CAP, including CO<sub>2</sub> transportation and storage, is 119.6 mills/kWh, a 59% increase over the base PC plant without CO<sub>2</sub> capture. The LCOE increase caused by CO<sub>2</sub> capture for the Hot-CAP is 31% lower than the benchmark MEA-based process.

## References

- [1] DOE/NETL DE-FOA-0000403. Attachment 3- Basis for Technology Feasibility Study. (1/31/2011)
- [2] The DOE/NETL report on “*Cost and Performance Baseline for Fossil Energy Plants, Volume 1: Bituminous Coal and Natural Gas to Electricity*” (Revision 2, November 2010), NETL Report No. DOE/NETL-2010/1397.
- [3] Chemstations, Inc. “ChemCad Version 6 User Guide”. (2012).  
[http://www.chemstations.com/content/documents/CHEMCAD\\_6\\_User\\_Guide\\_2012.pdf](http://www.chemstations.com/content/documents/CHEMCAD_6_User_Guide_2012.pdf)
- [4] Otimeas Treating, Inc. The Origins of ProTreat™. The Contactor. Vol 3 Supplement. October 2, 2009.
- [5] Nexant Inc. report on “Techno-Economic Study Results & Methodology for Post-Combustion Flue Gas CO<sub>2</sub> Capture Using Gas Pressurized Stripping Technology”. (October, 2012).

## Appendix - Acronyms and Abbreviations

AMP	2-Amino-2-methyl-1-propanol
Ar	Argon
BOP	Balance of Plant
CCS LLC	Carbon Capture Scientific, LLC
CO <sub>2</sub>	Carbon Dioxide
CT	Cooling Tower
CTB	Carbonate-To-Bicarbonate Conversion
CW	Cooling Water
DEA	Diethanolamine
DOE	US Department of Energy
FGD	Flue Gas Desulfurization
H <sub>2</sub> O	Water
Hg	Mercury
HHV	Higher Heating Value
Hot CAP	Hot Carbonate Absorption Process
HP	High Pressure
Hr, hr, h	Hour
IEP	Innovations for Existing Plants
ILP	Intermediate Low Pressure
K <sub>2</sub> CO <sub>3</sub>	Potassium Carbonate
KPa	Kilo Pascal for Pressure
kWe	Kilowatt electric
kWh	kilowatt hour
lb	Pound Mass
LCOE	Levelized Cost of Electricity
LP	Low Pressure
ME	Major Equipment
MEA	Monoethanolamine
mm	Milimeter
MM	million
MPa	Mega Pascal for Pressure
MWe	Megawatt electric
N <sub>2</sub>	Nitrogen
Na <sub>2</sub> CO <sub>3</sub>	Sodium Carbonate
NETL	National Energy Technology Laboratory
NO	Nitric Oxide

NO <sub>2</sub>	Nitrogen Dioxide
NO <sub>x</sub>	Oxides of Nitrogen
O&M	Operating and Maintenance
O <sub>2</sub>	Oxygen
PC	Pulverized coal
PCC	Post-Combustion Capture
PM	Particulate Matters
ppmv	Parts per Million by Volume
psig	Pounds Per Square Inch, gauge
PZ	Piperazine
Sat	Saturated
SCR	Selective Catalytic Reduction
SH	Superheat
SO <sub>2</sub>	Sulfur Dioxide
ST	Short Ton
STG	Steam Turbine Generator
TBD	To be determined
TDC	Total Direct Cost
TFC	Total Field Cost
TG	Turbine Generator
TPC	Total Plant Cost
TS&M	Transport, Storage & Monitoring
UIUC	University of Illinois at Urbana-Champaign
vol%	Percentage by Volume