Bench-Scale Development of a Hot Carbonate Absorption Process with Crystallization-Enabled High Pressure Stripping for Post-Combustion CO₂ Capture

–Preliminary Techno-Economic Study Results and Methodology

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Submitted to

January 29, 2014

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Abstract

This report summarizes the methodology and preliminary results of a techno-economic analysis on a hot carbonate absorption process (Hot-CAP) with crystallization-enabled high pressure stripping for post-combustion CO_2 capture (PCC). This analysis was based on the Hot-CAP that is fully integrated with a sub-critical 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 techno-economic analysis addressed 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 Hot-CAP is 611 MWe, greater than that with Econoamine (550 MWe). The total capital cost for the Hot-CAP, including CO_2 compression, is $399 million, less than that for the Econoamine PCC ($493 million). O&M costs for the power plant with Hot-CAP is $175 million annually, less than that with Econoamine ($178 million). The 20-year levelized cost of electricity (LCOE) for the power plant with Hot-CAP, including CO_2 transportation and storage, is 119.4 mills/kWh, a 59% increase over that for the plant without CO_2 capture. The LCOE increase caused by CO_2 capture for the Hot-CAP is 31% lower than that for its Econoamine counterpart.

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1 Introduction

The University of Illinois at Urbana-Champaign (UIUC) and Carbon Capture Scientific (CCS), LLC have been developing an absorption-based technology for post-combustion CO_2 capture (referred to as Hot-CAP). The Hot-CAP technology employs a carbonate salt, such as potassium carbonate (K_2CO_3) or sodium carbonate (Na_2CO_3), as a solvent. The process involves four major unit operations including CO_2 absorption into a high-concentration carbonate solution, crystallization of bicarbonate, bicarbonate slurry-based high pressure CO_2 stripping, and reclamation of a sulfate salt from the sulfur dioxide (SO_2) 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 Innovations for Existing Plants (IEP) 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 power plant equipped with the Hot-CAP technology for post-combustion CO_2 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.[1]

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[2]" was used as the main source of reference to be followed, as per the guidelines of ATTACHMENT 3 of DE-FOA-0000403. Carbon Capture Scientific, 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 pulverized coal-fired (PCF) power plant that utilizes an advanced Econoamine-based absorption system for CO_2 capture and compression.

For this techno-economic study, the Hot-CAP process replaces the Econoamine-based CO_2 absorption system in the original case. The objective of this study is to assess the performance of a full-scale Hot-CAP-based PCC design integrated with a subcritical PCF plant similar to Case 10 of the DOE/NETL report, such that it corresponds to a nominal 550 MWe subcritical PCF plant with 90% CO_2 capture. This plant has the same boiler firing rate and generated steam pressure as the DOE/NETL report's Case 10 PCF plant. However, due to the difference in performance between the Hot-CAP-based PCC and the Econoamine-based CO_2 absorption technology, the net power output of this plant may not be exactly at 550 MWe.

2 Design Basis

2.1 Power Plant Design Criteria

2.1.1 General

This study is based on a design of a subcritical PCF power plant with carbon capture to generate approximately a nominal net 550 MWe, consistent with the DOE/NETL-2010/1397 report's Case 10. The original gross output of the power plant is about 673 MWe. The steam generator for the subcritical PCF plants is a drum-type, wall-fired, balanced draft, natural circulation, totally enclosed dry bottom furnace, with superheater, reheater, economizer and air-heater. The steam turbine generator (STG) operates at throttle conditions of 16.5 MPa/566°C/566°C (2,400 psig/1,050°F/1,050°F).

The plant is designed for NO_x reduction using a combination of low-NOx burner and overfire air as well as with the installation of a selective catalytic reduction (SCR) system. Particulate control is designed with fabric filter/baghouse, which consists of two separate single-stage, in-line, multi-compartment units. Flue gas desulfurization (FGD) system is a wet limestone forced oxidation positive pressure absorber non-reheat unit, with wet-stack, and gypsum production for SO_2 removal. The combination of pollution control technologies used in the PCF plants, SCR, fabric filters and FGD, result in significant co-benefit capture of mercury. The mercury co-benefit capture is assumed to be 90% for this combination, sufficient to meet current mercury emissions limits, and hence no activated carbon injection is needed in this case.

The power plant is considered to operate as a base-loaded unit but with consideration for daily or weekly cycling. Annual capacity factor is 85% or 7,450 hrs/year at full capacity.

2.1.2 Site-Related Conditions

The subcritical PCF plant in this study is assumed to be located at a generic plant site in Midwestern USA, with site-related conditions as shown below:

- Location   Midwestern USA
- Elevation, ft above sea level   0
- Topography   Level
- Size, acres   300
- Transportation   Rail
- Ash/slag disposal   Off site
- Water   Municipal (50%)/Groundwater (50%)
- Access   Landlocked, having access by train and highway
- CO_2 disposition   Compressed to 152 bar at battery limit

2.1.3 Meteorological Data

Maximum design ambient conditions for material balances, thermal efficiencies, system design and equipment sizing are:

- Atmospheric pressure, kPa   101

- Dry bulb temperature, °C   15
- 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 PCF 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 PCF plant aux loads include cooling water (CW) circulation pump loads, cooling tower (CT) fan loads, and transformer loss. PCF 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_2 capture and compression processes, plus any new CW and CT consumptions due to the PCC cooling loads.

   Table 2.2. PCC-independent PCF 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
Total	31,170

- It is assumed that the subcritical PCF 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:
        Blowdown Losses = Evaporative Losses/(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_2	0.008
NOx	0.339
Particulate Matter (Filterable)	0.063
Hg	5.53E-6

Emission component NO_2, and SO_2 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 PCF power plant to capture up to 90% of the CO_2 in the flue gas. It is assumed that all of the fuel carbon is converted to CO_2 in the flue gas. CO_2 is also generated from limestone in the FGD system, and 90% of the total CO_2 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 PCF plant boiler will be burning 278,956 kg/hr of as-received Illinois No. 6 coal to generate 2364,000 kg/hr of SH HP steam to the STG based on Case 10 subcritical PCF 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_2 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_2	vol%	67.94
O_2	vol%	2.38
CO_2	vol%	13.50
Ar	vol%	0.81
H_2O	vol%	15.37
Total		100

2.2.3 Design CO2 Product Specifications

Recovered CO2 is delivered at the battery limit with the following specifications:
- Inlet pressure, MPa   15.3
- Inlet temperature, °C   26
- N_2 + Ar concentration, ppmv   < 1000 (revised for PCC processes)
- O_2 concentration, ppmv   < 100 (revised for PCC processes)
- H2O, 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 desuperheated to 10 °C above saturation temperature to allow positive control of desuperheater condensate injection. Degree of ILP steam superheat can be varied to meet minimum desuperheater design requirement.

- Return Condensate
  Reboiler steam condensate will be pumped back to the power plant hot at:
        Minimum pressure, MPa   1.2
        Temperature, °C   to be determined (TBD) by PCC Design

- 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 ℃ and atmospheric pressure, where CO_2 and other acid gases are absorbed into a carbonate (K_2CO_3 or PC) solution. The CO_2-rich carbonate solution exiting the absorption column is cooled through a cross-flow heat exchanger by the CO_2-lean carbonate solution returning from the crystallization tank. After passing the cross-flow heat exchanger, the CO_2-rich carbonate solution enters to the crystallization tank, where potassium bicarbonate ($KHCO_3$) salt crystals are formed due to the low solubility of the bicarbonate at low temperatures (30-40℃). 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 ℃). The CO_2 stream released in the stripper contains a relatively small amount of water vapor. The CO_2-rich gas stream exiting the stripper is further cooled, dehydrated, and compressed to a sequestration-ready pressure. The CO_2-lean solution exiting the bottom of the stripper enters the crystallization tank after exchanging heat with the feed slurry.
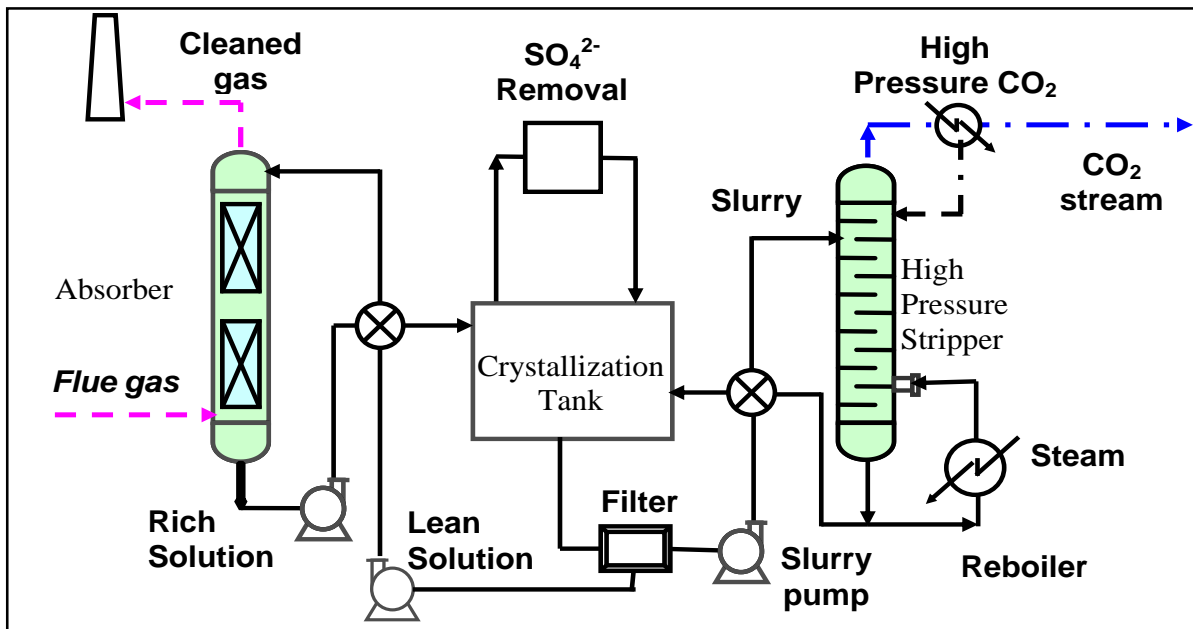

Figure 3.1. Schematic diagram of the proposed Hot-CAP

The composition of the CO_2-rich stream from the absorption column is 35~40 wt% (K_2CO_3-equivalent) PC solution with about 50% carbonate-to-bicarbonate (CTB) conversion. After the $KHCO_3$ crystallization, the CTB conversion level of the lean stream is 20% or less, which returns to the absorption column. The concentration of $KHCO_3$ in the absorption and crystallization process is subject to its solubility under different conditions as shown in Figures 3.2.
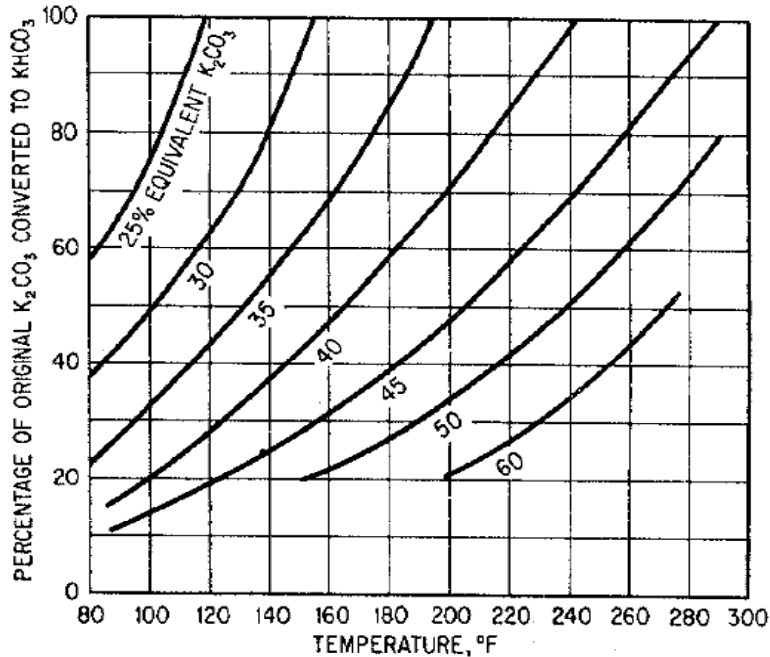
Figure 3.2. Effects of temperature and CTB conversion on the solubility of $KHCO_3$ in $K_2CO_3/KHCO_3$ solutions

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, Carbon Capture Scientific 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.
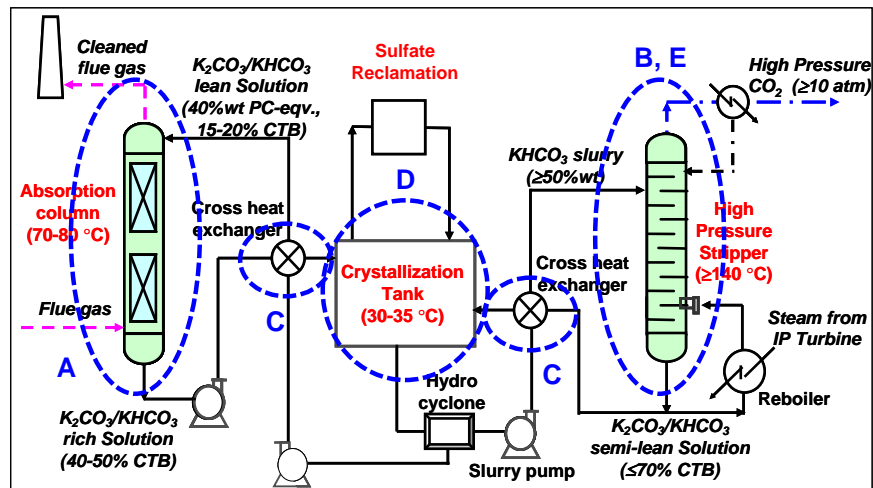
Figure 3.3. Five major technical risks identified for the Hot-CAP.

Risk A is related to the rate of CO_2 absorption at elevated temperature (60-80 °C) and concentration of PC 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

Risk	Mitigation	Risk ID
Rate of CO_2 absorption at temperature (60°C-80°C) and concentration of PC solution (~40wt.%) 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 $\geq$ 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 University of Illinois at Urbana-Champaign to screen promoters that can accelerate the rate of absorption. The promoters were evaluated by the measured CO_2 removal efficiency by the promoted 40 wt% PC solution in an absorption column. The selected experimental results are illustrated in Figure 3.4**.**
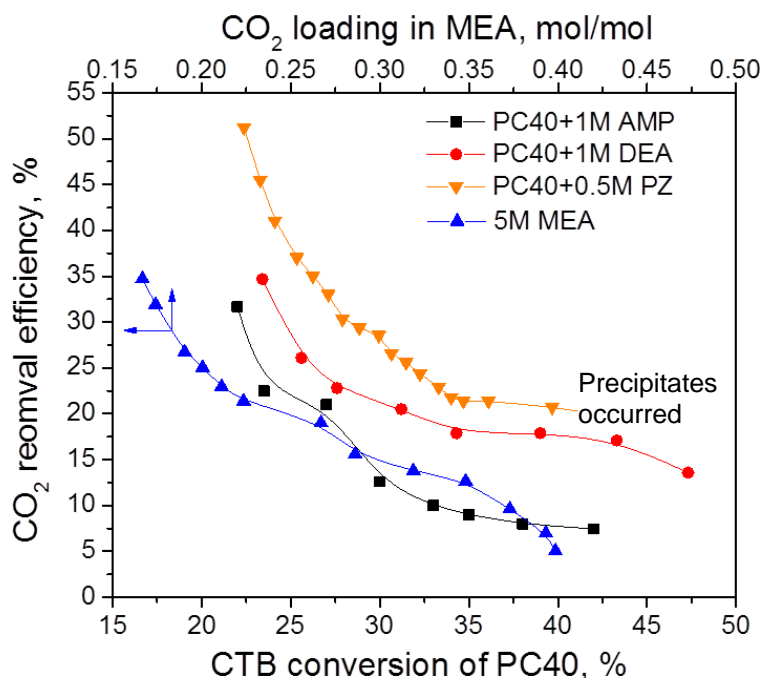
Figure 3.4. Experimental results of CO_2 absorption performance in the 40 wt% PC solution with the addition of various promoters.

Details of the experimental study was described in the previous quarterly reports (e.g., for October 2012 to December 2012). The key results from the experimental study include:
- CO_2 removal efficiency was low in the absence of a promoter
- CO_2 removal efficiency by the 40 wt% PC promoted with 1M DEA or 0.5M PZ, either for the CO_2 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_2 absorption into PC solutions. Both thermodynamic and kinetic behaviors of the CO_2 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 power 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_2 absorption into PC solutions at 70 °C are summarized in Figures 3.5, 3.6, 3.7 and 3.8. The results demonstrate that the CO_2 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_2 into the 40 wt% PC solution with a CO_2 loading equivalent to 15% initial CTB conversion is able to achieve the targeting 90% CO_2 removal at L/G ratios above 7. However, the absorption of CO_2 into the PC with 20% initial CTB conversion cannot achieve the targeting 90% CO_2 removal within a reasonable range of L/G ratios. In addition, a high PC concentration is favorable for the CO_2 removal.

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- The kinetic analysis for the CO_2 absorption into PC 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 PC 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% PC with 20% initial CTB conversion promoted by 0.5M PZ at 60 °C.
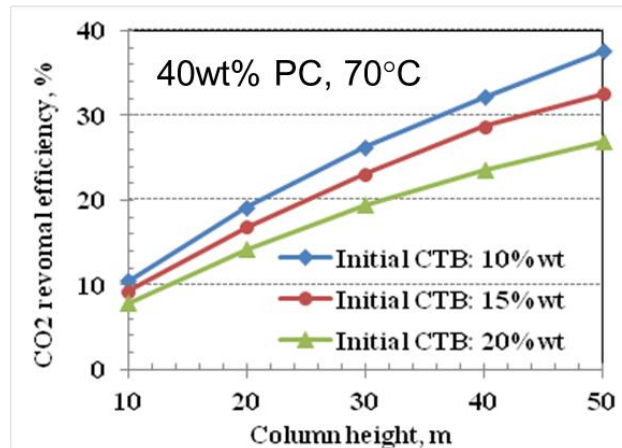
Figure 3.5 Simulation results of CO_2 absorption into the 40 wt% PC solution without a promoter.
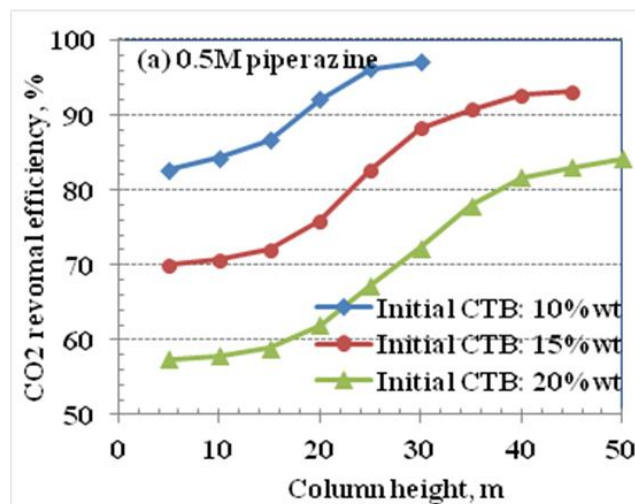
Figure 3.6. Simulation results of CO_2 absorption into the 40 wt% PC with the addition of 0.5M PZ as a promoter.
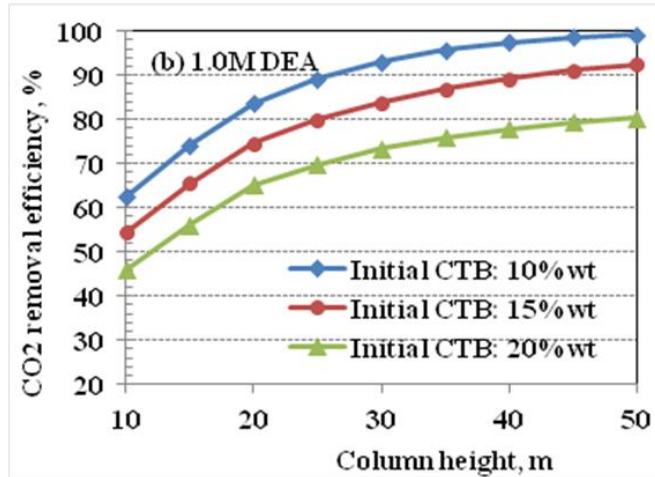
Figure 3.7. Simulation results of CO_2 absorption into the 40 wt% PC with the addition of 1.0M DEA as a promoter.
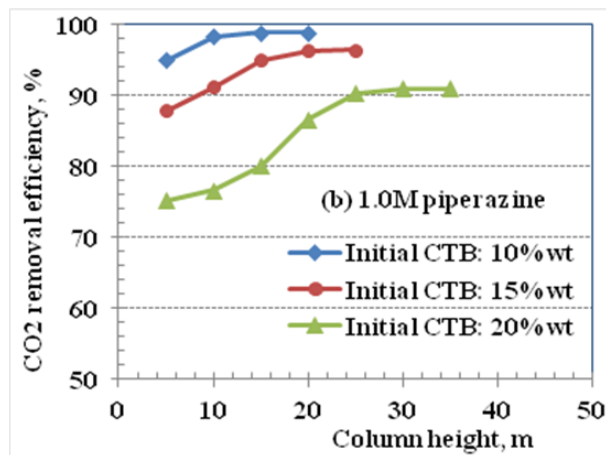
Figure 3.8. Simulation results of CO_2 absorption into the 40 wt% PC with the addition of 1.0M PZ as a promoter.

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_2 stripping in a high pressure stripping column. The stripper design is critical since a high stripping pressure and a smaller water vapor/CO_2 partial pressure ratio will significantly reduce the stripping heat (associated with water vaporization) during CO_2 stripping and the required compression work downstream. The measured VLE data confirmed the feasibility of high pressure CO_2 stripping in the Hot-CAP process. A higher stripping temperature, a higher level of CTB conversion, and a higher PC concentration led to a higher stripping pressure and a lower water vapor/CO_2 ratio. However, recent studies at Carbon Capture Scientific LLC revealed that it is not economically beneficial for the CO_2 stripping operating at extraordinarily high pressures because of the following two reasons:

- High temperature steam required by a high pressure stripping reduces the net electric power generation;

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- Extra power is consumed for pumping the circulation solvent to an extraordinarily high pressure.

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:
- Reducing the temperature difference between the streams in the cross-flow heat exchangers
- Pre-seeding of the crystallization solution
- Using plate and frame type of heat exchangers
- Using a vacuum cooling crystallizer or a surface cooling crystallizer equipped with scrappers
- 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 PC 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 assisting in 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 Risk *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 (i.e. bicarbonate slurry) and high CO_2 loading (high bicarbonate/carbonate ratio) for the regenerated lean solution. On the other hand, higher CO_2 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 a power plant steam at a relatively high

temperature (about 180 °C) as a heat source for the stripper reboiler.

The results from the above risk analysis are considered in the sequent techno-economic studies. These results are especially important affecting the equipment capital cost. The use of a series of CSTRs in the crystallizer design will have a large impact on the overall capital costs of the Hot-CAP process.
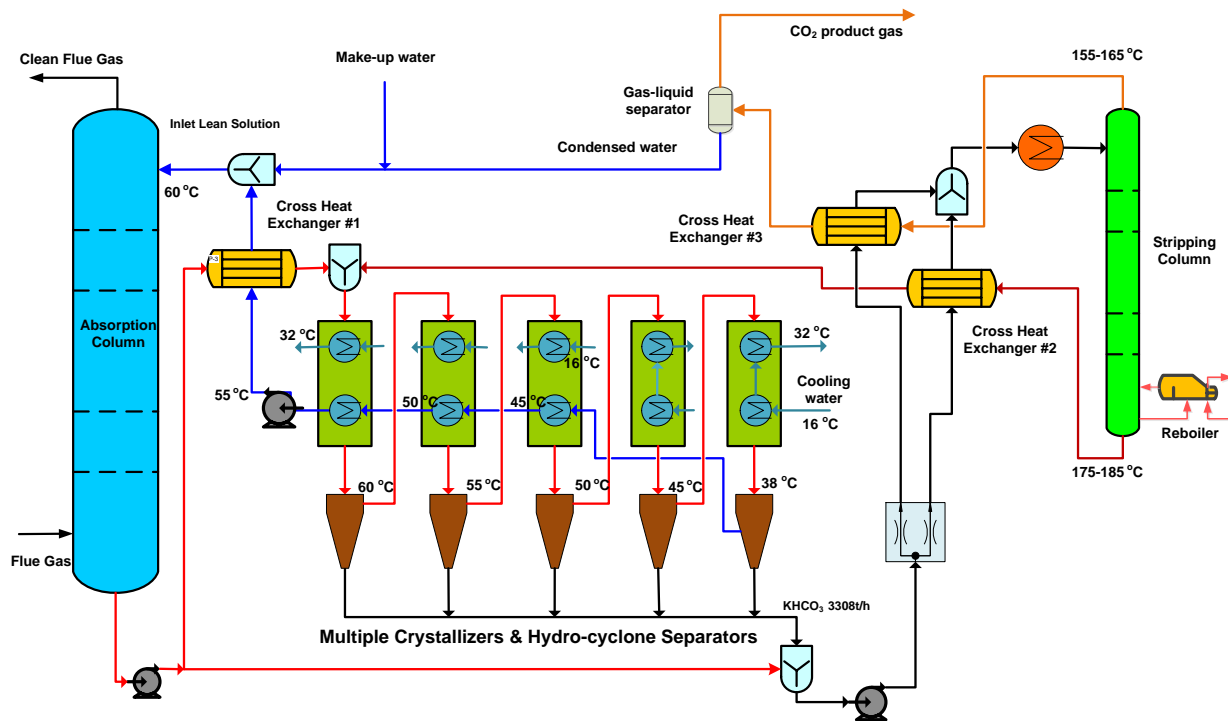

Figure 3.9. A multiple crystallizer unit design developed to address Risk D.

3.3 Design of Hot-CAP process

3.3.1 Design of Absorption Column

In addition to the above risk analysis, process simulations using ProTreat Program for the absorption column has resulted in the following recommendations:

- A promoter, either DEA or PZ, is required to achieve the targeting 90% CO_2 removal. PZ is recommended because it can be used at a high temperature, i.e. 160 °C, without encountering significant degradation.
- A more concentrated PC solution is preferred for achieving a large CO_2 working capacity. However, the PC concentration will be limited by the solubility of $KHCO_3$ in the rich solvent. A compromised PC concentration used in this study is 40 wt% K_2CO_3-equivalent.
- The initial level of CTB conversion (i.e., lean CO_2 loading) was selected by a tradeoff between the absorption and stripping performance. 20% initial CTB conversion is applied in this study.
- The temperature of the inlet lean solvent has a great effect on that of the effluent rich solvent. A higher inlet temperature will result in a larger heat loss caused by evaporating

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water carried out in the purified flue gas thus reducing the temperature of the rich solvent. 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_2 recovery while 0.5M PZ is sufficient to achieve 90% CO_2 recovery at 60 °C. On the other hand, a low inlet solvent temperature will significantly reduce the solubility of $KHCO_3$ in the rich solvent, which in turn will reduce the solvent's working capacity. Therefore, the inlet temperature of the lean solvent is adopted at 60 °C.

In summary, the hot carbonate solvent is a 40 wt% (K_2CO_3-equivalent) PC solution with 20% initial CTB conversion containing 0.5M PZ as a promoter. The design temperature of the inlet lean solvent is 60 °C. Because the absorption in the design solvent solution is sufficiently fast, the required flow rate of the solvent is limited by the $KHCO_3$ solubility in the rich solvent. Simulation results using ProTreat Program show that the outlet temperature of the rich solvent reaches 67 °C, at which the solubility of $KHCO_3$ 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 size of absorption column required for the targeting 90% CO_2 recovery is 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 cannot provide sufficiently accurate performance predictions when the regeneration temperature in the stripping column 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_2 compression work. Therefore, the simulation of stripping column is based on a self-developed steady-state thermodynamic model using the vapor-liquid-equilibrium (VLE) data measured in this project. The experimental VLE data for CO_2 and water vapor are shown in Figures 3.10 and 3.11.
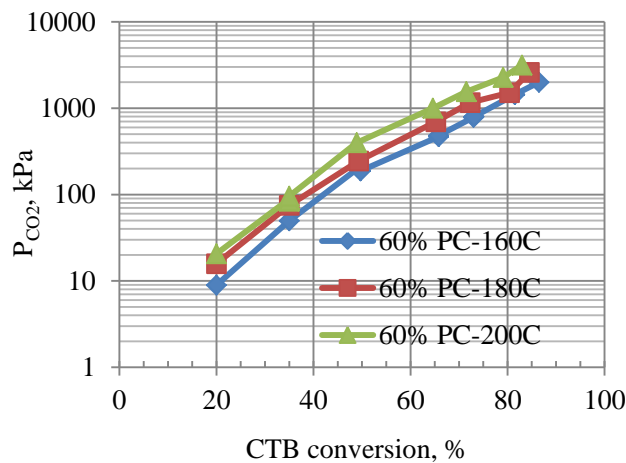
Figure 3.10. Experimental VLE data for CO_2 in 60 wt% K_2CO_3 equivalent solution
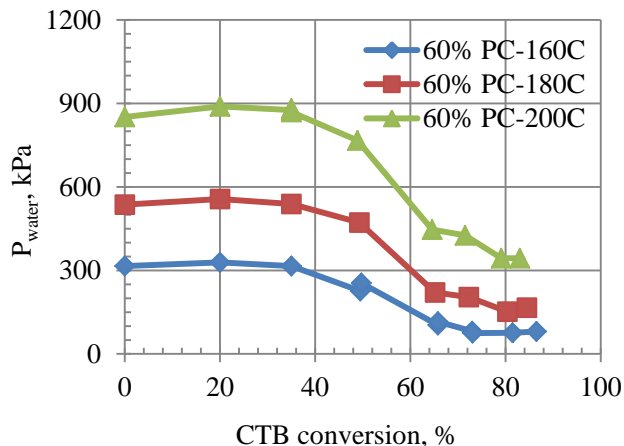
Figure 3.11. Experimental VLE data for water vapor in 60 wt% K_2CO_3 equivalent solution

In the simulation, the following assumptions and designs are made to simplify 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) PC 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 the 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 Software, the temperature difference between the top rich solution and the bottom lean solution is usually between 17and 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 reboiler at the bottom of stripper 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 Software. As the VLE database built in ProTreat 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 of reduced to 120 °C, simulation results showed that the 29% CTB conversion in the hot

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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_2 reaction kinetics usually increases with increasing, the striping 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 $KHCO_3$ solids in the inlet rich solution (79% CTB conversion and 161 °C) entering of 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 hydro-cyclone. 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 $KHCO_3$ crystallization is instantaneous but it takes about a 30 minute residence time 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, 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 $KHCO_3$ 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 °C to 32 °C was adopted to determine the cooling water flow rate required for the 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 $W/m^2 \cdot K$.

Table 3.2 Estimation of volumes for five stages of crystallizers

Stage of Crystallizer	1st	2nd	3rd	4th	5th	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^3	309	355	293	155	269	1381
Design crystallizer volume, m^3	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 $KHCO_3$ salt: 81% of the rich solution from the absorber (16,142 tonne/hr of 40 wt% PC with 45% CTB conversion at 67 °C) and all of the lean solution regenerated from the stripper (6,406 tonne/hr of 60 wt% PC with 29% CTB conversion at 70 °C). The density of the blended solution is estimated at 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 $KHCO_3$ crystallization heat. The rate of $KHCO_3$ crystallization in each crystallizer was estimated based on its solubility as shown in Figure 3.2. The total $KHCO_3$ crystallization heat released was estimated based on the crystallization rate and the molar crystallization heat (26.2 kJ/mole $KHCO_3$). The sensible heat was estimated using ProTreat Software. Table 3.2 summarizes the design parameters of the five crystallizers. It should be noted that the cooling load in the first crystallizer is the smallest because there was no crystallization and thus no related heat release at this stage because the crystal was pre-formed during the heat exchange for the both sources of solutions. The total volume of the five crystallizers amount to 3,220 m^3, which results in an actual total residence time 15 minutes. Thus, such a residence time is sufficient to meet both the heat exchange and crystal growth requirement. The total cooling water load reached 393.5 MW, requiring a cooling water flow of 21,183 tonne/hr.
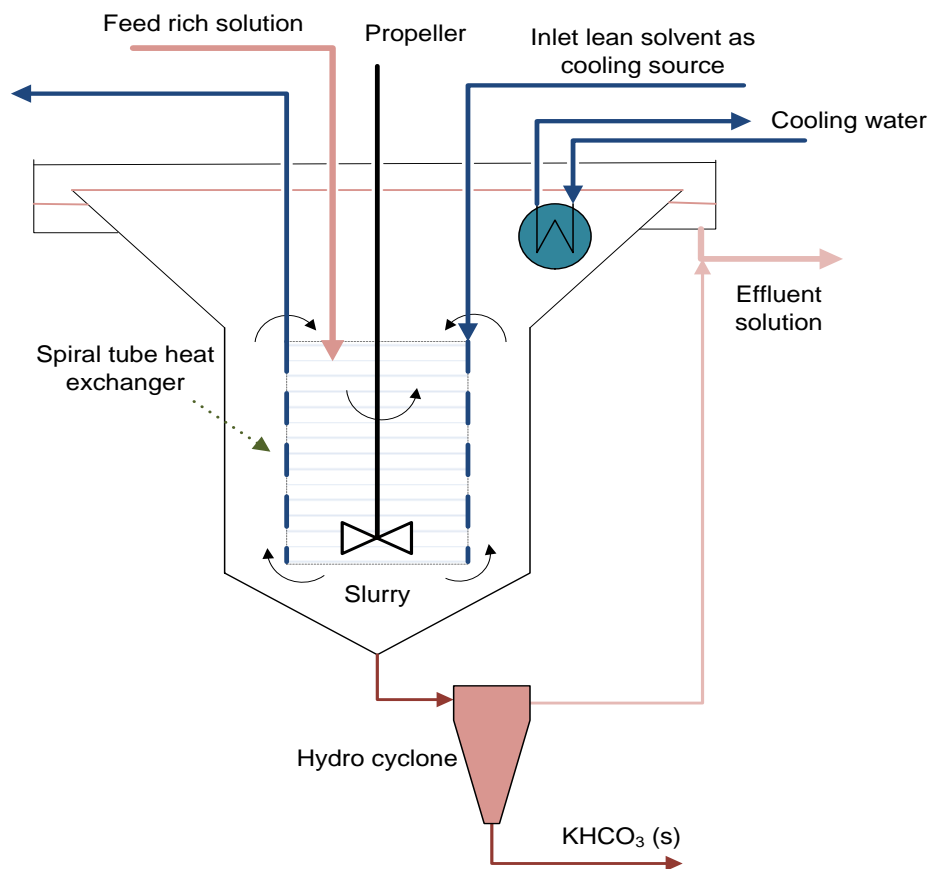
Figure 3.12. A schematic of crystallizer structure.

3.4 Mass and Heat Balance of Hot-CAP Process

Based on the composition and flow rate of the flue gas specified in Case 10 in the DOE/NETL-2010/1397 report, the results of mass and heat balances for the Hot-CAP-based PCC are shown in Figure 3.13. The absolvent is a 40 wt% PC solution with 20% lean CTB conversion containing 0.5M PZ at a total flow rate of 19,300 tonne/hr. 90% CO_2 removal is achieved in the absorption column, where the CO_2 loading of the solvent increases from 20% (lean) to about 45% (rich) CTB conversion. The rich PC 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) PC slurry with 79% CTB conversion.

In the crystallization process, 3,308 tonne of $KHCO_3$ solids are obtained from the blended solution made of 81% of the 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% PC solution of 20% CTB conversion as the lean solution returning to the absorber. During the crystallization, external cooling water is required to control the temperatures of the different crystallizer units.

Although the crystallizers are designed to use covered tanks, water losses during the crystallization process are 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 temperature (60 °C) at the absorber inlet. Totally 598 tonne/hr of CO_2 is released from the regeneration of $KHCO_3$ during the stripping process. Afterheat recovery from the regenerated hot lean solvent and the CO_2 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 the stripping column. The total heat consumption in the Hot-CAP is estimated at 2,382 kJ/kg CO_2.
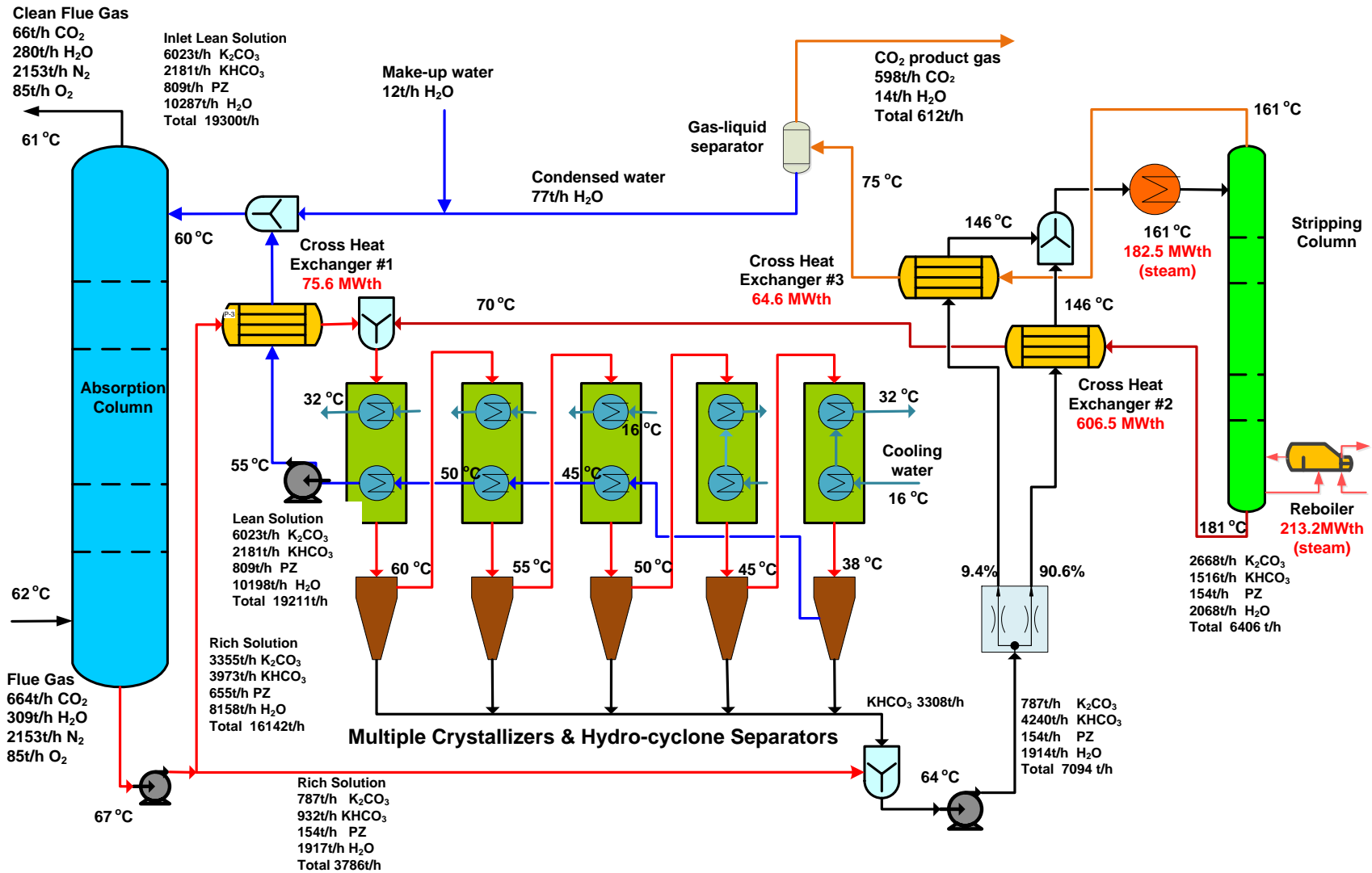
Figure 3.13. A schematic flowchart of the Hot-CAP process with mass and heat balances.

Clean Flue Gas
66t/h CO_2
280t/h H_2O
2153t/h N_2
85t/h O_2

Inlet Lean Solution
6023t/h K_2CO_3
2181t/h KHCO$_3$
809t/h PZ
10287t/h H_2O
Total 19300t/h

Make-up water
12t/h H_2O

CO$_2$ product gas
598t/h CO_2
14t/h H_2O
Total 612t/h

161 °C

Gas-liquid separator

Condensed water
77t/h H_2O

75 °C

61 °C

60 °C

146 °C

161 °C
182.5 MWth (steam)

Stripping Column

Cross Heat Exchanger #1
75.6 MWth

70 °C

Cross Heat Exchanger #3
64.6 MWth

146 °C

P-3

32 °C

16 °C

32 °C

Cross Heat Exchanger #2
606.5 MWth

55 °C

50 °C

45 °C

Cooling water
16 °C

Lean Solution
6023t/h K_2CO_3
2181t/h KHCO$_3$
809t/h PZ
10198t/h H_2O
Total 19211t/h

Absorption Column

62 °C

60 °C

55 °C

50 °C

45 °C

38 °C

9.4%

90.6%

181 °C

2668t/h K_2CO_3
1516t/h KHCO$_3$
154t/h PZ
2068t/h H_2O
Total 6406 t/h

Rich Solution
3355t/h K_2CO_3
3973t/h KHCO$_3$
655t/h PZ
8158t/h H_2O
Total 16142t/h

Multiple Crystallizers & Hydro-cyclone Separators

KHCO$_3$ 3308t/h

787t/h K_2CO_3
4240t/h KHCO$_3$
154t/h PZ
1914t/h H_2O
Total 7094 t/h

Reboiler 213.2MWth (steam)

Flue Gas
664t/h CO_2
309t/h H_2O
2153t/h N_2
85t/h O_2

67 °C

Rich Solution
787t/h K_2CO_3
932t/h KHCO$_3$
154t/h PZ
1917t/h H_2O
Total 3786t/h

64 °C

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4 Cost Estimation

4.1 Cost Estimation Methodology

The Total Plant Cost (TPC) and Operation & Maintenance (O&M) costs for the subcritical PCF power plant and the equipped CO_2 capture plant are estimated based on 2007 costs, using methodology introduced from ATTACHMENT 3 of the DOE/NETL DE-FOA-0000403[1].

4.1.1 Capital Cost

The DOE/NETL-2010/1397 report provided a cost estimate for 14 major subsystems of the Case 10 subcritical PCF plant with CO_2 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 PCF 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 PCF plant subsystems and bases for modifications are shown in Table 4.1.

- Subcritical PCF Plant

The capital cost estimates for the subcritical PCF of the overall 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_2 capture section provided in the report, resulting in a variation of the PCF plant performance due to the differences in PCC design as well as solvent selection. As stated in Section 2.1.4, the revised PCF plant with Hot-CAP-based PCC performance 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 PCF plant. For this reason, the PCF 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 and 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 PCF 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 ±30%. Separate estimates are prepared for the CO_2 recovery facility and the CO_2 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 total plant cost (TPC), the Engineering and Construction Management Fees and Home office cost, and contingencies are added to the TFC to come up with the TPC.

Table 4.1. Cost estimate basis for subcritical PCF plant with CO_2 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	PCF BOILER	Capacity Factor	AR Coal
5	FLUE GAS CLEANUP	Capacity Factor	AR Coal
5B	CO2 REMOVAL & COMPRESSION		
5B.1	CO2 Removal System	New estimate	N/A
5B.2	CO2 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 sizes of absorption columns and stripping columns are determined in the previous section. The size for each piece of other major equipment used in the process is estimated based on the individual heat and material stream flows of the simulation.

When the sizes for the individual equipment except for crystallization tanks were determined, the costs for the equipment and installation labor were estimated by scaling up from the cost estimates for corresponding equipment pieces in a techno-economic analysis for "Post-Combustion Flue Gas CO_2 Capture Using Gas Pressurized Stripping Technology" made by Nexant Inc for Carbon Capture Scientific LLC. The costs of crystallization tanks were estimated based on literature references and quotes obtained from vendors.

Costs for bulk materials such as instrumentations, piping, structure steel, insulation, electrical, painting, concrete & 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' report.

- 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 plants over their expected life. These costs include:
- o Operating labor
- o Maintenance – material and labor
- o Administrative and support labor
- o Consumables
- o Fuel
- o 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 economic analysis for this study follows the same methodology as used in 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 PCF plant without CO_2 capture (Case 9) LCOE is to be used as the benchmark for the subcritical PCF plant with CO_2 capture to compare against. The Case 9 20-year 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 reboiling steam requirement and electrical consumption. The process' steam consumption is used to estimate the gross power generated by the power plant's steam turbines.

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	1.3MPa		Load	Circ. rate	
		kW	tonne/hr	tonne/hr	tonne/hr	MW	tonne/hr	MW
	Exchangers							
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	
	Pumps and drivers							
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						
	Compressors							
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_2 compression section utilities

Item No.	Item Name	Elec. Power	Steam	Water Condensate	Cooling Water		Regeneration duty
			0.8MPa		Load	Circ. rate	
		kW	tonne/hr	Kg/hr	MW	tonne/hr	MW
	Exchangers						
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	
	Compressors						
K-101	Stage 1 CO_2 compressor	9,717					
K-102	Stage 2 CO_2 compressor	9,444					
K-103	Stage 3 CO_2 compressor	8.956					
K-104	Stage 4 CO_2 compressor	7,007					
K-105	Stage 4 CO_2 compressor	2,849					
	Packaged Equipment						
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

The auxiliary loads for the overall plant are separated into three categories: PCC-independent PCF plant auxiliary loads, PCC-dependent PCF auxiliary loads, and PCC loads. The PCC-independent PCF 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 PCF aux loads, 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 PCF auxiliary loads. Tables 4.2 and 4.3 show the utility consumption of the Hot-CAP-based PCC plant's CO_2 capture section and CO_2 compression section, respectively.

4.4 Capital Cost Estimate for Hot-CAP PCC

The Hot-CAP-based process' CO_2 capture and CO_2 compression section major equipment (ME) lists are provided in Tables 4.4 and 4.5, respectively. The estimated total field costs (TFC) for the Hot-CAP-based PCC CO_2 capture section (totaling $226.2 million) and CO_2 compression section (totaling $36.5 million), which include 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 at $262.7 million, is listed in Table 4.8.

Table 4.4. Hot-CAP-based CO_2 capture section major equipment list

VESSELS & TANKS:

Item No.	Item Name	Type	Design Conditions kPa	Design Conditions °C	Material of Construction	Quantity per Lot	Units	Inside Diameter m	Ht or Tan/Tan Length m	Width m	Length m	Number of Lots	2007 Equip Cost $1000
C-101	Absorption Column	Vert	108	70	Kill CS	1	Vessel	14.8	24.0			2	25,253
	- Sulzerpak	250.0 Sulzerl			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 Sulzerl			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	Horizont	600	75	304 SS	1	Vessel	2.3	2.6			1	119
C-104	Stg 1 Crystallizer	Horizont	101	60	Concrete	1	Tank	5.0	9.6			1	129
C-105	Stg 2 Crystallizer	Horizont	101	55	Concrete	1	Tank	5.0	11.7			1	193
C-106	Stg 3 Crystallizer	Horizont	101	50	Concrete	1	Tank	5.0	12.8			1	229
C-107	Stg 4 Crystallizer	Horizont	101	45	Concrete	1	Tank	5.0	13.9			1	271
C-108	Stg 5 Crystallizer	Horizont	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:

Item No.	Item Name	Type	Design P, kPa Shell	Design P, kPa Tube	Des Temp, °C Shell	Des Temp, °C Tube	Material Of Construction Shell	Material Of Construction Tube	Duty MW	Total Bare Tube Area, sq. m	In Series	In Parallel	Total # Req	Total Equip Cost $1000
E-100	Absorber R/L HX #1	P&F	800	800	190	190		304SS	75.6	3086	1	3	3	370
E-101	Stripper bottom R/L HX #2	P&F	800	800	190	190		304SS	606.5	13433	1	10	10	1,610
E-102	CO2 product gas HX #3	P&F	800	800	190	190		304SS	64.6	1125	1	1	1	135
E-103	Stripper inflow heater	P&F	800	800	190	190		304SS	182.5	3408	1	3	3	409
E-104	Stripper reboiler	S&T	1300	800	190	190		304SS	213.2	13325	1	1	1	3,199
E-105	Stage 1 crystallizer cooler by CW	STE		400		190		304SS	14.1	283	1	1	1	34
E-106	Stage 2 crystallizer cooler by CW	STE		400		190		304SS	70.4	1632	1	2	2	196
E-107	Stage 3 crystallizer cooler by CW	STE		400		190		304SS	26.1	713	1	1	1	85
E-108	Stage 4 crystallizer cooler by CW	STE		400		190		304SS	123.6	4123	1	6	6	494
E-109	Stage 5 crystallizer cooler by CW	STE		400		190		304SS	159.3	7165	1	10	10	859
E-110	Stage 1 crystallizer cooler by Mother	STE		400		190		304SS	74.5	7962	1	10	10	954
E-111	Stage 2 crystallizer cooler by Mother	STE		400		190		304SS	73.4	7846	1	10	10	941
E-112	Stage 3 crystallizer cooler by Mother	STE		400		190		304SS	101.7	7109	1	10	10	852
												Sum	68	10,138

COMPRESSORS, BLOWERS & DRIVERS:

Item No.	Item Name	Type	Design Conditions kPa	Design Conditions °C	Wheel or Impel'r	Casing	Des Flow SCMH	Inlet kPa	Delta P kPa	Comp BHP	HP	Type	Total # Req	Total Equip Cost $1000
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:

Item No.	Item Name	Type	Design Conditions kPa	Design Conditions °C	Wheel or Impel'r	Casing	Des Flow LPS	Inlet kPa	Delta P kPa	Pump BHP	HP	Type	Total # Req	Total Equip Cost $1000
G-101	Absorber bottom rich Pump	Cent.	300	70	CS	CS	3543	101	200	494	519	Motor	8	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

Item No.	Item Name	Type	Tube Design Cond kPa	Tube Design Cond °C	Mat Of Constr	Insulation Thickness Inches	Ht, m	Width, m	Total Length, m	Total # Req	Total Equip Cost $1000
L-100	Flue Gas Feed & Exhaust Ducts	Duct	103	60	CS	1	4.5	4.5	1262		16731
										Sum	16731
										TOTAL EQUIP COST	
											72417

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

VESSELS & TANKS:

Item No.	Item Name	Type	Design Conditions kPa	Design Conditions °C	Material of Construction	Quantity per Lot	Units	Inside Diameter m	Ht or Tan/Tan Length m	Width m	Length m	Number of Lots	Total Equip Cost $1000
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 Shell	Design P, kPa Tube	Des Temp, °C Shell	Des Temp, °C Tube	Material Of Construction Shell	Material Of Construction Tube	Duty MW	Total Bare Tube Area, sq. m	Physical Arrangement In Series	Physical Arrangement In Parallel	Physical Arrangement Total # Req	Total Equip Cost $1000
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 kPa	Design Conditions °C	Material Of Construction Wheel or Impel'r	Material Of Construction Casing	Design Capacity Des Flow SCMH	Design Capacity Inlet kPa	Design Capacity Delta P kPa	Design Capacity Comp BHP	Driver HP	Driver Type	Total # Req	Total Equip Cost $1000
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 kPa	Design Conditions °C	Material Of Construction Wheel or Impel'r	Material Of Construction Casing	Design Capacity Des Flow LPS	Design Capacity Inlet kPa	Design Capacity Delta P kPa	Design Capacity Pump BHP	Driver HP	Driver Type	Total # Req	Total Equip Cost $1000
													Sum	

PACKAGED & MISC EQUIPMENT:

Item No.	Item Name	Type	Tube Design Cond kPa	Tube Design Cond °C	Mat Of Construct	Design Capacity	Remarks	Total # Req	Total Equip Cost $1000
V-100	TEG Dehydration Package	Pkg						1	1722
								Sum	1722
								TOTAL EQUIP COST	**13788**

Table 4.6. Hot-CAP-based PCC CO$_2$ capture section total field cost

COST CODE	DESCRIPTION	QTY	MEAS Unit	UNIT COSTS MATL	LABOR	SC/Other	D HIRE UNIT MH	TOTAL MHRS * S/C	D HIRE	Equipment	BULK	LABOR	SC/Other	TOTAL
	PROCESS EQUIPMENT & DUCTWORK													
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													
	SITEWORK													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$_2$ compression section total field cost

COST CODE	DESCRIPTION	QTY	MEAS Unit	UNIT COSTS MATL	UNIT COSTS LABOR	UNIT COSTS SC/Other	D HIRE UNIT MH	TOTAL MHRS * S/C	TOTAL MHRS * D HIRE	COSTS IN U.S.$1000 Equipment	COSTS IN U.S.$1000 BULK	COSTS IN U.S.$1000 LABOR	COSTS IN U.S.$1000 SC/Other	TOTAL
	PROCESS EQUIPMENT & DUCTWORK													
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
	SITEWORK													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
	TOTAL (2007 BASIS)													36,498

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Table 4.8. Hot-CAP-based PCC total field cost

COST CODE	DESCRIPTION	QTY	MEAS Unit	UNIT COSTS MATL	UNIT COSTS LABOR	UNIT COSTS SC/Other	D HIRE UNIT MH	TOTAL MHRS * S/C	TOTAL MHRS * D HIRE	COSTS IN U.S.$1000 Equipment	COSTS IN U.S.$1000 BULK	COSTS IN U.S.$1000 LABOR	COSTS IN U.S.$1000 SC/Other	TOTAL
	PROCESS EQUIPMENT & DUCTWORK													
	Hot CAP-based CO2 CAPTURE TRAIN 1	1	Train							37,535	8,588	26,294		72,417
	Hot CAP-based CO2 COMPRESSION TRAIN 1	1	Train							12,317		1,472		13,789
	FREIGHT	5.00	%							2,493				2,493
	TOTAL PROCESS EQUIPMENT & DUCTWORK		EA							52,345	8,588	27,765		88,699
	INSTRUMENTS													12,053
	PIPING													42,258
	STEELWORK													6,580
	INSULATION													5,252
	ELECTRICAL													23,375
	CONCRETE													9,302
	BUILDING													906
	SITEWORK													10,672
	PAINTING													590
	TOTAL OTHER DIRECT COSTS													110,988
	SUBTOTAL DIRECT COSTS													199,686
	SUBTOTAL CONSTRUCTION INDIRECT COSTS													62,982
	SUBTOTAL FIELD COSTS (2007 BASIS)													262,669
	TOTAL (2007 BASIS)													262,669

4.5 Performance Summary of Subcritical PCF Plant with Hot-CAP PCC

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

Plant Performance Summary*	Case 10 with Econamine PCC	Case 10 with Hot-CAP PCC	
Consumables:			
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	
PCC Steam Extraction:		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*
Power Generation Summary, kW:			
Steam Turbine Gross Power	672,700	*722,695*	
Auxiliary Load Summary, kW:			
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_2 Capture Plant Auxiliaries	22,400	*24,070*	
CO_2 Compression	48,790	*39,307*	
Total Auxiliaries, kW	**122,740**	**111,381**	
Net Power Export, kW	**549,960**	**611,314**	
Net Plant Efficiency, % HHV	**26.2**	***29.1***	
Net Plant Heat Rate, kJ/kWh	**13,764**	***12,404***	
Cooling Tower Loads, GJ/hr:			
Surface Condenser Duty	2,034	*2,670*	
CO_2 Capture Plant Cooling Duties	3,585	*1,417*	
CO_2 Compression Cooling Duties	Incl Above	*278*	
Total Cooling Tower Loads, GJ/hr	**5,619**	***4,365***	
Overall Makeup Water Balance, m³/min:			
FGD Makeup	5.5	5.5	
BFW Makeup	0.4	0.4	
CO_2 Capture & Compression Makeups	0.2	*0.2*	
Cooling Tower Makeup	36.4	*28.3*	
Net Raw Water Makeup, m³/min	**42.5**	***34.4***	

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 PCF plant (Case 10 of the DOE/NETL report). Because the condition and amount of steam requirement for this Hot-CAP-based PCC plant is different from those with Econoamine-based PCC in Case 10, there are certain differences between this plant and the Case 10 subcritical PCF plant with the Econoamine-based PCC:

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

Therefore, the net power output and thermal efficiency of the subcritical PCF plant with Hot-CAP-based CO_2 capture is different from those of the Case 10. Table 4.9 summarizes the performance and thermal efficiency of the overall PCF plant with Hot-CAP-based PCC and compares to those for the subcritical PCF plant with Econoamine-based PCC. The net power output and efficiency of the subcritical PCF plant with Hot-CAP CO_2 capture are 611 MWe and 29.1%, higher than corresponding 550 MWe and 26.2% for the sub-critical PCF plant with Econoamine-based CO_2 capture.

4.6 Capital Cost Estimate for Subcritical PCF Plant with Hot-CAP PCC

Based on the cost estimating methodology for the overall subcritical PCF 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 PCF Plant with PCC capital cost. The total plant cost for the PCF plant with Hot-CAP is $1,523 million.

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

											Contingency		
Acct No.	Plant Description	Capacity Scaling Units	Capacity	Equip Cost	Material Cost	Direct Labor Cost	Indirect Labor Cost	Sales Tax	Bare Erect Cost	Eng'g CM H.O. & Fee	Process	Project	Total Plt Cost
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^3/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	HRSG, 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^3/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
	TOTAL COST			587,715	141,757	448,663			1,178,135	110,959	46,235	187,325	1,522,653

Hot CAP PCC Total Plant Cost Details in U.S.$1000

4.7 O&M Cost Estimate for Subcritical PCF Plant with Hot-CAP PCC

According to Section 4.1, the annual O&M costs consist 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.

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

Fixed Operating Expenses:			Expense
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
Total Fixed Operating Costs			**$50,607,198**
Variable Operating Expenses:			
Maintenance Material Cost			$14,982,908
Consumables	Consumption/day	Unit Cost	
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_2CO_3 Solvent (ton)	7.72	725.60	$1,737,906
NaOH (ton)	7.89	433.68	$1,061,704
H_2SO_4 (ton)	7.53	138.78	$324,217
Corrosion Inhibitor	0	0	$0
Act Carbon (lb)	168	1.05	$54,728
Ammonia (19% NH3)(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
Total Variable Operating Costs			**$124,506,733**
TOTAL ANNUAL OPERATING COSTS			**$175,113,930**

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 fuel, 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. Because half of the net raw water makeup comes from groundwater, water expenditure was calculated based on another half of the net raw water makeup from the overall plant water balance. PCF 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 potassium carbonate solvent. The potassium carbonate solvent replacement cost is estimated based on the total degradation rate caused by the reaction with SO_2. For this study, it is assumed that deep FGD is included in the PCC so that the total degradation follows the DOE/NETL Case 10 sulfur balance. Corrosion inhibitor is not required due to the assumption that Hot-CAP is less corrosive than MEA.

The total estimated annual O&M costs for the subcritical PCF plant with Hot-CAP-based PCC is $175 million. Table 4.11 shows the breakdown of the O&M costs in a similar format to the DOE/NETL Case 10 report.

4.8 LCOE Estimate for Subcritical PCF Plant with Hot-CAP PCC

Using the methodology as described in Section 4.1, the estimated LCOE for the subcritical PCF 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 PCF plant is 119.4 mills/kWh, which is 159% of the LCOE (75.3 mills/kWh) of the same PCF plant without CO_2 capture. Compared to a 139.0 mills/kWh LCOE for the MEA PCC-based PCF plant, the increase of LCOE caused by CO_2 capture with the Hot-CAP is 31% lower than that with its Econoamine counterpart.

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

Post-Combustion Case Description	Subcritical PCF w/o CO2 Capture	Subcritical PCF w/ CO2 Capture	
Case Number	**Case 9**	**Case 10**	**This Study**
Type of CO$_2$ Capture Technology	**N/A**	**Econoamine**	**Hot CAP**
Capital Cost Year	2007	2007	2007
CO$_2$ Capture	0%	90%	90%
Power Production, MW			
Gross Power	583	673	723
Net Power	550	550	611
Cost			
Total Plant Cost, 2007$/kW	1,662	2,942	2,491
Total Overnight Cost, 2007$/kW	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
Total Overnight Cost, 2007$ ×1000	1,098,124	1,985,432	1,865,250
Total As Spent Capital, 2007$/kW	2,264	4,115	3,478
COE (mills/kWh, 2007$)	59.4	109.6	94.1
CO$_2$ TS&M Costs	0.0	5.8	5.2
Fuel Costs	15.2	21.3	19.2
Variable costs	5.1	9.2	8.1
Fixed Costs	7.8	13.1	11.1
Capital Costs	31.2	60.2	50.4
LCOE, mills/kWh	**75.3**	**139.0**	**119.4**
% of Case 9 LCOE	**100%**	**185%**	**159%**

5 Summary

A techno-economic analysis was performed to compare the energy use and cost performance of a nominal 550 MWe subcritical PCF plant without CO_2 capture (DOE/NETL Case 9), with the Econoamine-based PCC (DOE/NETL Case 10), or with the Hot-CAP-based PCC. The results show that the net power produced in the PCF plant equipped with Hot-CAP is 611 MWe, greater than that with Econoamine (550 MWe) mainly because the Hot-CAP uses a lesser amount of steam for CO_2 stripping.

The Hot-CAP has a slightly lower capital cost ($48 million) in the CO_2 compression section than the Econoamine PCC ($50 million) because the stripped product gas has a higher pressure (6 verse 2 bar). The Hot-CAP also has a lower capital cost in the CO_2 capture section ($351 million) than the Econoamine PCC ($443 million) because it requires a smaller stripping column and doesn't need a polishing unit for deep sulfur removal.

O&M costs for the PCF plant equipped with Hot-CAP are estimated to be $175 million annually, less than that of the PCF plant with Econoamine ($178 million).

The 20-year LCOE for the PCF plant with Hot-CAP, including CO_2 transportation and storage, is 119.4 mills/kWh, a 59% increase over that for the PCF plant without CO_2 capture. The LCOE increase caused by CO_2 capture for the Hot-CAP is 31% lower than that for its Econoamine counterpart.

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

[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_2 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	Carbon Capture Scientific, LLC
CO_2	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_2O	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_2CO_3	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_2	Nitrogen
Na_2CO_3	Sodium Carbonate
NETL	National Energy Technology Laboratory
NO	Nitric Oxide

NO$_2$	Nitrogen Dioxide
NOx	Oxides of Nitrogen
O&M	Operating and Maintenance
O$_2$	Oxygen
PCF	Pulverized coal-fired
PC	Potassium Carbonate
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$_2$	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