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# CaO-based CO<sub>2</sub> Capture Technology and Its Application in Power Plants

Rongrong Zhai and Yongping Yang  
*North China Electric Power University  
China*

## 1. Introduction

Carbon dioxide (CO<sub>2</sub>) emission causes many environmental problems, such as global warming, which results in the rise of sea level. Currently, it is a mission for the whole world to control and reduce the emission of CO<sub>2</sub>. There is a great amount of CO<sub>2</sub> being released from coal-fired power plants. Coal is expected to continue to be a prominent fuel for electricity production in the future (IEA, 2003; Thitalamol et al., 2007). CO<sub>2</sub> mitigation options include energy efficiency improvements, the switch to less carbon-intensive fuels, nuclear power, renewable energy sources, enhancement of biological sinks, reduction of non-CO<sub>2</sub> greenhouse gas emissions and CO<sub>2</sub> capture and storage (CCS). CCS has the potential to reduce overall mitigation costs and increase flexibility in achieving greenhouse gas emission reductions. The widespread application of CCS would depend on technical maturity, costs, overall potential, diffusion and transfer of the technology to developing countries and their capacity to apply the technology, regulatory aspects, environmental issues and public perception (IPCC, 2005).

In order to mitigate the CO<sub>2</sub> emission from coal-fired power plants, there are three main methods of CCS technologies such as post-combustion capture, pre-combustion capture and oxy-fuel combustion. Among them, the post-combustion method is the easiest method for the retrofitting of the current existing power plants. Limestone has been viewed as a potential sorbent for CO<sub>2</sub> capture process because of its low-cost and easy access. The basic process of CO<sub>2</sub> capture systems using lime was first outlined by Shimizu et al. (1999).

## 2. Related work

### 2.1 Modeling of CaO based CO<sub>2</sub> capture process

There have been extensive researches on the post-combustion systems incorporated with carbonation/calcinations looping cycle, involving the reaction kinetics, modeling and so on. Microscopic modeling mainly focuses on the particle or a small group of particles. The random pore model (RPM) (Khinast et al., 1996; Adanez et al., 2000; Stanmore and Gilot, 2005), grain model (Garcia-Labiano et al., 2002; Stanmore and Gilot, 2005; Garea et al., 2005) and homogeneous particle model (Garcia-Labiano et al., 2002; Abanades et al., 2004) all belong to this catalogue. On the other hand, the macroscopic modeling focuses on the reactors where those processes take place. The core reactors used for calcination and

carbonation are either bubbling fluidized bed reactors or circulating fluidized bed reactors. It is therefore necessary to examine the models of those reactors in order to analyze the lime based CO<sub>2</sub> capture system. The first approximation to the modeling of a fluidized bed carbonator reactor has been proposed and the efficiency of CO<sub>2</sub> capture is analyzed (Alonso et al., 2009). The KL fluid bed model (Kunii and Levenspiel, 1990) has been used to interpret the experimental CO<sub>2</sub> concentration profiles measured inside the bed during the fast reaction period in literatures (Abanades et al., 2004; Romano, 2009).

## **2.2 Integration of CaO based CO<sub>2</sub> capture process with power plants**

Experiments in a pilot-scale fluidized-bed reactor have been carried out to investigate the carbonation reaction of CaO, as a potential method for CO<sub>2</sub> capture from combustion flue gases at high-temperatures. The heat requirements for calcium looping cycle have been developed (Rodriguez et al., 2008). The carbonate looping processes of post-combustion CO<sub>2</sub> capture for coal-fired power plants have been studied (Strohle et al., 2009; Hawthorne et al., 2009). The carbonate looping process was integrated after the flue gas desulphurization unit of an existing power plant. The energy penalty is about 2.75%, much lower than that of other CO<sub>2</sub> capture technologies (Strohle et al., 2009). The retrofitted steam cycle which includes an air separation unit (ASU) and integrated CO<sub>2</sub> conditioning unit with interstage cooling increases the net generated power from 1052 MWe to 1533 MWe, resulting in an overall electric efficiency of 39.2% (Hawthorne et al., 2009). One kind of possible integrations of a power plant with CaO-based CO<sub>2</sub> capture process has been studied by Romeo et al. (2008).

## **2.3 Contributions of our work**

For the modeling of carbonator, the parameters that may affect the carbonation process haven't been fully analyzed in the previous research. We built the model of carbonator which can be used to examine the effects of the parameters on the carbonation process. For the integrations of CO<sub>2</sub> capture process and power plants, few papers have been focused on the effects of the CaO-based CO<sub>2</sub> capture process to the reference plant which vents its CO<sub>2</sub> to the atmosphere. We examined the possible integrations of the power plant with the CO<sub>2</sub> capture process.

In this chapter, a one-dimension model of CaO-based CO<sub>2</sub> capture process has been built. Based on this model, the effects of superficial gas velocity, the particle diameter and the calcinations/carbonation cycle on the carbonation process have been discussed. In addition, five cases for the possible integration with the reference plant which vents its CO<sub>2</sub> to the atmosphere are considered. Three possible methods for the utilization of the recovered heat from the CO<sub>2</sub> capture process are used in these integrations. In addition, thermal performance evaluation of these cases has been investigated.

## **3. Carbon capture process based on CaO**

### **3.1 System description**

CO<sub>2</sub> capture plant as shown in Fig.1 is composed of two fluidized bed reactors. One reactor is the calciner, operating at about 900 °C, using oxygen obtained from the air separation unit (ASU); the other one is the carbonator, operating at around 650 °C. For the simplicity of the study, sulphation has been taken as undesirable reaction and been neglected in our analysis.

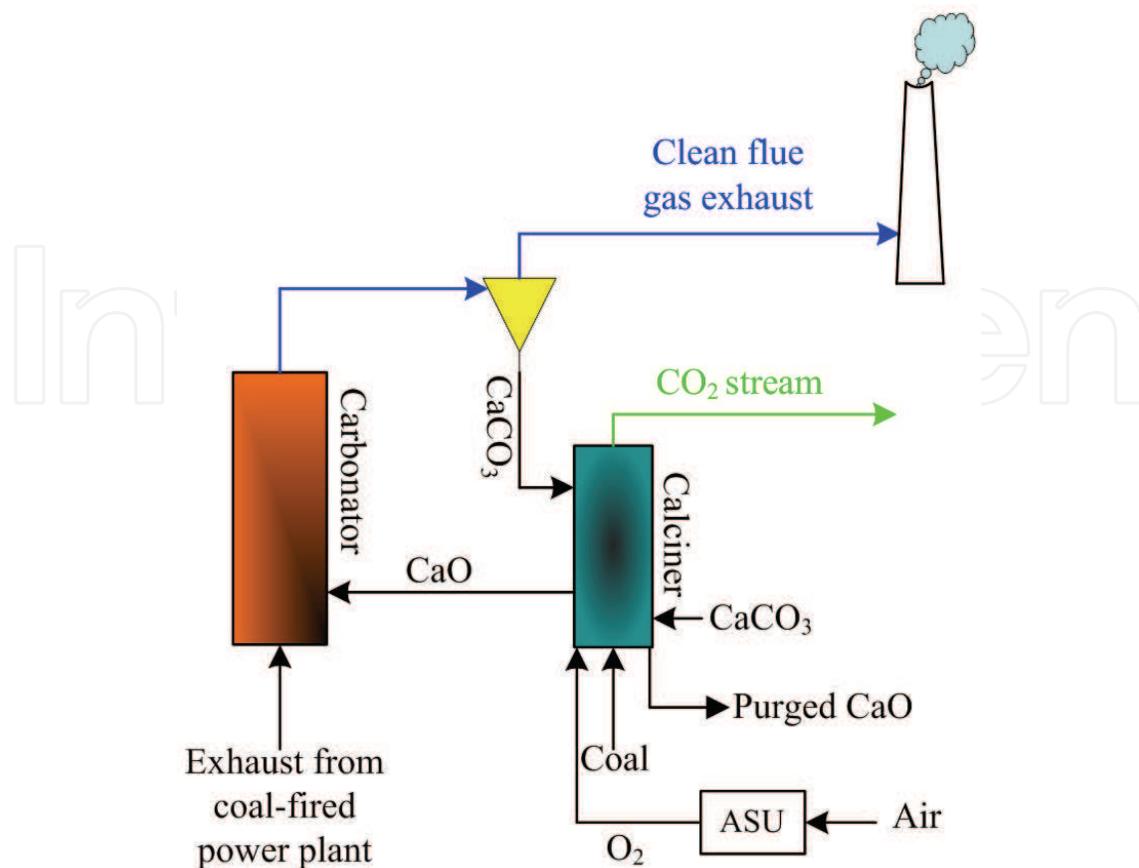
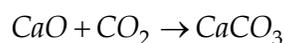
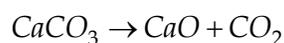


Fig. 1. General layout of CaO-based CO<sub>2</sub> capture cycle

In the carbonator the main reaction is:



In the calciner the main reaction is:



Pipes and cyclones are used to exchange particles between the calciner and carbonator. Ideally, the process can take place continuously without refilling fresh limestone. However, the reactivity of the sorbent decays as the process goes on. Therefore, used lime needs to be purged out of the carbonator and new lime needs to be added to maintain these reactions.

### 3.2 Modeling of the carbonator

The carbonator as shown in Fig.1 is one key reactor in the CO<sub>2</sub> capture process, where the carbonation takes place. The dense bed height of the carbonator is about 1.5m. In order to interpret the carbonation process, the model of the carbonator has been built. It is composed of three sub-models: the CO<sub>2</sub> mass balance sub-model, the hydrodynamics sub-model and the microscope carbonation sub-model. The CO<sub>2</sub> mass balance sub-model (KL fluid bed model) was first proposed by (Kunii and Levenspiel, 1990), and then it has been used in literatures (Abanades et al., 2004; Romano, 2009). The main equations for hydrodynamics sub-model are semi-experimental equations from previous research. The microscopic carbonation sub-model describes the carbonation rate of the process.

### The CO<sub>2</sub> mass balance sub-model

Assumptions have been made for the CO<sub>2</sub> mass balance sub-model. It is assumed that there are two phases in the reactor as the bubble phase and the emulsion phase. The carbonation reaction takes place only in the emulsion phase, assuming that the bubble phase is particle free.

Equilibrium CO<sub>2</sub> concentration in the carbonation/calcination process  $C_{CO_2,eq}$  at about 650°C was given by (Baker, 1962).

$$C_{CO_2,eq} = \frac{1.46210^{11}}{T} \exp(-19130 / T) \quad (1)$$

Where, T is the temperature of carbonation.

CO<sub>2</sub> mass balance in bubble phase (Kunii and Levenspiel, 1990):

$$-u_b \frac{dC_{bCO_2}}{dz} = \gamma_b f_a K_r (C_{bCO_2} - C_{CO_2,eq}) + K_{be} (C_{bCO_2} - C_{eCO_2}) \quad (2)$$

CO<sub>2</sub> mass balance in emulsion phase (Kunii and Levenspiel, 1990):

$$-(1 - \delta) u_{mf} \frac{dC_{eCO_2}}{dz} = (1 - \delta)(1 - \varepsilon_{mf}) f_a K_r (C_{eCO_2} - C_{CO_2,eq}) - \delta K_{be} (C_{bCO_2} - C_{eCO_2}) \quad (3)$$

Where,  $C_{bCO_2}$  is CO<sub>2</sub> concentration in the bubble phase;  $C_{eCO_2}$  is CO<sub>2</sub> concentration in the emulsion phase;  $u_{mf}$  is Minimum fluidization velocity;  $\varepsilon_{mf}$  is Bed porosity at minimum fluidization;  $\delta$  is Bubble fraction in the fluidized bed;  $\gamma_b$  is Volume of solids in bubble phase divided by the volume of bubbles.

### The hydrodynamics sub-model

As mentioned before, the hydrodynamics sub-model has been built based on previous semi-experimental equations. The main factors to describe the hydrodynamics are minimum fluidization velocity, the equivalent volume diameter, the absolute velocity of the bubble, bubble fraction in the fluidized bed and effective gas velocity in bubble phase.

The minimum fluidization velocity can be defined as the velocity at which the bed just becomes fluidized. The minimum fluidization velocity can be obtained by solving the following equations (Grace, 1992; Basu, 2006):

$$Re_{mf} = \frac{U_{mf} d_p \rho_g}{\mu} = [27.2^2 + 0.0408 Ar]^{0.5} - 27.2 \quad (4)$$

The bubble diameter  $d_b$  can be estimated using the expression given by Darton et al. (Darton et al., 1977):

$$d_b = 0.54(U - U_{mf})^{0.4} (Z + 4\sqrt{A_0})^{0.8} g^{-0.2} \quad (5)$$

The absolute velocity of the bubble  $U_b$  has been developed by Davidson and Harrison (Davidson and Harrison, 1963) as

$$U_b = (U - U_{mf}) + 0.711\sqrt{gd_b} \quad (6)$$

Bubble fraction in the fluidized bed  $\delta$  (Kunii and Levenspiel, 1990):

$$\delta = \frac{u_0 - u_{mf}}{u_b + 2u_{mf}} \quad (7)$$

Effective gas velocity in bubble phase  $u_b^*$  (Kunii and Levenspiel, 1990):

$$u_b^* = \frac{u_0 - (1 - \delta)u_{mf}}{\delta} \quad (8)$$

$$u_b^* = u_b + 3u_{mf} \quad (9)$$

### The microscopic carbonation sub-model

In the microscopic carbonation sub-model, the key parameters are: the overall gas interchange coefficient between bubble and emulsion phases, carbonation reaction rate constant, overall carbonation rate constant of particles in the emulsion phase, fraction of active CaO in the carbonator and maximum carbonation conversion of CaO to CaCO<sub>3</sub> in the N<sup>th</sup> cycle.

Overall gas interchange coefficient between bubble and emulsion phases  $K_{be}$  (Kunii and Levenspiel, 1990):

$$K_{be} = 4.5u_{mf} / d_b \quad (10)$$

Carbonation reaction rate constant  $K_{ri}$  (Kunii and Levenspiel, 1990):

$$K_{ri} = k_s \frac{X_{b,N} S_0 \rho_{CaO}}{M_{CaO}} (1 - X)^{2/3} \quad (11)$$

The correlation for the Sherwood number  $Sh$  (Turnbull and Davidson, 1984) has been listed as Eq.12, in which the CO<sub>2</sub> mass transfer coefficient towards the carbonating particles  $k_g$  can be estimated.

$$Sh = \frac{D_{CO_2}}{k_g d_p} = 2\varepsilon_{mf} + 0.95 Re_{mf}^{0.5} Sc^{0.3} \quad (12)$$

Overall carbonation rate constant of particles in the emulsion phase  $K_r$  (Abanades et al., 2004):

$$K_r = \frac{1}{\frac{d_p}{6k_g} + \frac{1}{K_{ri}}} \quad (13)$$

Fraction of active CaO in the carbonator  $f_a$ :

$$f_a = X_{b,N} - X \quad (14)$$

With the development of research, the equation for the maximum carbonation conversion of CaO to CaCO<sub>3</sub> in the N<sup>th</sup> cycle has been improved.

Firstly, Abanades (Abanades, 2002) fitted an empirical model:

$$X_N = f^{N+1} + b \quad (15)$$

Where  $f=0.782$  and  $b=0.174$  are constant.

Then, Abanades and Alvarez (Abanades and Alvarez, 2003) developed an analytical model in terms of change in porosity, as:

$$X_{b,N} = f_m^N (1 - f_w) + f_w \quad (16)$$

Where,  $f_m=0.77$ ,  $f_w = 0.17$ .

Later, Wang and Anthony (Wang and Anthony, 2005) presented a model based on only one parameter:

$$X_N = \frac{1}{1 + kN} \quad (17)$$

Finally, considering a residual conversion,  $X_r$ , of about 7-8%, Grasa and Abanades (Grasa and Abanades, 2006) proposed a semi-empirical model based on one parameter:

$$X_N = \frac{1}{\frac{1}{1 - X_r} + kN} + X_r \quad (18)$$

$N$  is defined as the number of calcination/carbonation cycle, with  $N=1$  the first carbonation reaction using fresh CaO. The fresh CaO is obtained from the calcination of limestone.

## 4. Integration and evaluation of a power plant with a CaO based CO<sub>2</sub> capture process

### 4.1 System description

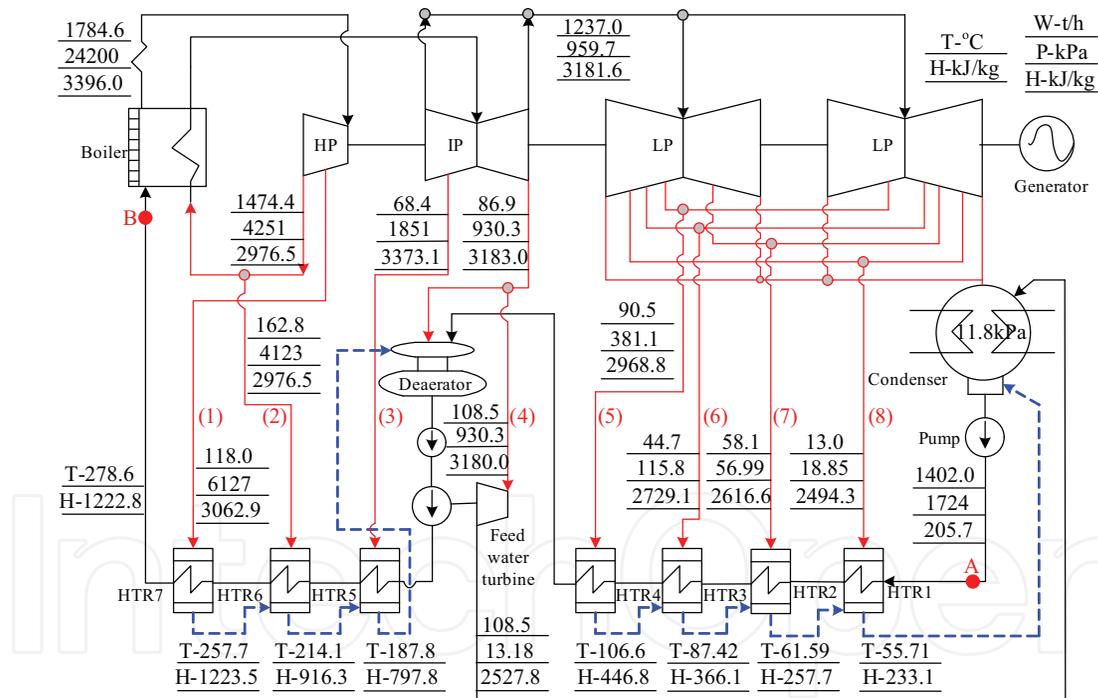
The basic layout of CaO-based CO<sub>2</sub> capture process has been shown in Fig. 1. Table 1 shows the main variable of the capture plant.

Oxygen mass flow (kg/s)	78.5
CaCO <sub>3</sub> mass flow (kg/s)	89.7
Coal mass flow (kg/s)	38.4
Coal composition (% wet basis)	
Carbon	66.9
Ash	13.9
Hydrogen	3.7
Nitrogen	1.6
Oxygen	6.5
Sulphur	0.7
Moisture	6.8
LHV(kJ/kg)	25200
Purge mass flow (kg/s)	64.5
Low CO <sub>2</sub> concentration flue gases (kg/s)	540.1
CO <sub>2</sub> to compression, captured (kg/s)	230

Table 1. Capture process main variables

Gross power output (MW)	600
Flue gases mass flow (kg/s)	647.6
Flue gases composition (% vol)	
CO <sub>2</sub>	12
N <sub>2</sub>	73.7
H <sub>2</sub> O	8.3
O <sub>2</sub>	5.5
SO <sub>2</sub>	0.5
Coal composition (% wet basis)	
Carbon	42.2
Ash	23.5
Hydrogen	2.7
Nitrogen	0.7
Oxygen	7.0
Sulphur	4.8
Moisture	19.1
LHV(kJ/kg)	15854

Table 2. Main operating conditions of Reference plant which vents its CO<sub>2</sub> to the atmosphere



\* (1)-(8) denote the stream conditions.

Fig. 2. Steam layout of a 600MW supercritical power plant

The main operating conditions of the existing power plant are shown in Table 2 with the typical composition of flue gas and coal composition referred to that in the literature (Romeo et al., 2008). The typical widely used 600MW reheated steam cycle system in recent power industry has been shown in Fig. 2. The combustion of coal takes place in the boiler. The unsaturated boiler feedwater from the condenser enters into the boiler after going through four low-pressure reheaters (HTR1, HTR2, HTR3, HTR4), three high pressure

reheaters (HTR5, HTR6, HTR7) and a deaerator (Deaerator). The outlet superheated steam from the boiler is transported to the high pressure cylinder to produce power, and then the exhaust steam drives intermediate pressure and lower pressure cylinders after being reheated in the boiler. In the end, the final exhaust is condensed in the condenser. It can be seen from Figure 2 that the extractions from different positions of the cylinders ((1)-(8)) are used to heat the feedwater via feedwater reheaters.

#### 4.2 Integration methods

The carbonation process is exothermic, i.e. a great amount of heat will be released during this reaction. It can be seen from Figure 1 that there are mainly three parts of recoverable heat from the CO<sub>2</sub> capture process as heat from the carbonator (Q<sub>1</sub>), heat from 650 °C clean flue gas (Q<sub>2</sub>) and heat from 900 °C CO<sub>2</sub> stream (Q<sub>3</sub>). Theoretically, these three parts of heat can be recovered and utilized. In the carbonator, Q<sub>1</sub> is related to three items: reaction heat released from the carbonation with about 168.5kJ per mol CaCO<sub>3</sub> produced; heat recovery from the 900 °C solid particles of 1227kg/s and heat consumed by heating the flue gas to 650 °C. Q<sub>1</sub> can be calculated by adding the first two items and deducting the last item. The maximum Q<sub>1</sub> is calculated to be about 389MW with flue gas temperature of 150 °C.

There are three possible methods for the utilization of the recovered heat from the CO<sub>2</sub> capture process, as follows (Yang et al., 2010):

1. Use the recovered heat to produce steam in order to replace the extracted steam from turbines (as shown (1)-(8) in Figure 2). The produced steam will be used to provide heat for the high-pressure heaters, deaerator or low-pressure heaters as the original extracted steam does;
2. Use the recovered heat to provide part of the boiler heat load;
3. Use the recovered heat to produce superheated steam, which can be then used to generate electricity in the turbines.

Based on the above three methods of heat utilization, five possible cases have been considered to integrate the CO<sub>2</sub> capture process with the power plant.

The basic heat transfer equation used in the calculation is as follows:

$$Q = C_p M (T_2 - T_1) \quad (19)$$

Where, C<sub>p</sub> is the specific heat capacity, depending on the temperature, pressure and composition of the gas. Once the temperature, pressure and composition of the gas are defined, C<sub>p</sub> can be calculated using MTDATA software which predicts the phases forming at equilibrium in systems containing many components and many phases.

##### Case 1: Use the recovered heat to replace extracted steam from the turbines

It can be seen from Figure 2 that there are eight steam extraction streams from turbines ((1)-(8)) used to heat the feedwater. In Case 1, the feedwater is initially pressurized via a series of pumps to the defined pressure values of the original extractions from the turbines. If the total mass flow of the extracted steam from the turbines decreases, there will be more electricity generated.

In Case 1, the condensed water from Point A (as seen in Figure 2) is divided into eight streams respectively. The recovered heat from the 650 °C clean flue gas is used to heat six feedwater streams; while the recovered heat from the 900 °C CO<sub>2</sub> stream is used to heat two feedwater streams. The total recovered heat from the CO<sub>2</sub> capture process is about 498.3MW with 342.3MW from the 650 °C clean flue gas and 156MW from the 900 °C CO<sub>2</sub> stream. The parameters of the steam heating system in Case 1 are the same with those in Figure 2. The

total increased power output for the power system due to having less steam extractions is about 148.8MW.

**Case 2: Use the recovered heat to replace part of the boiler heat load**

It has been calculated that the recovered heat can account for about 50% of the boiler load. However, it should be pointed out that the quantity of fuel for the boiler, the CO<sub>2</sub> releasing and the operating load for the CO<sub>2</sub> capture process will all reduce with part of the boiler's load replaced by the recovered heat. Therefore, the above factors should be considered when designing the heat recovery method in this case study.

It is assumed that the load of the CO<sub>2</sub> capture process is linear with the CO<sub>2</sub> emission. The heat replacement ratio is defined as:

$$X = \frac{Q_1 + Q_2 + Q_3}{Q_1 + Q_2 + Q_3 + Q_0} \quad (20)$$

Where  $Q_1$  is recoverable heat from the carbonator;

$Q_2$  is recoverable heat from the clean flue gas;

$Q_3$  is recoverable heat from CO<sub>2</sub> stream;

$Q_0$  is heat consumed by the boiler in the original steam system;

Based on the data from Table 1 & 2, the heat replacement ratio for the system is about 43.1% if all recovered heat is used to assume part of the boiler's load.

In Case 2, the 650 °C clean flue gas is used to heat the feedwater of 438.6t/h from 278.6 °C to 566 °C, while its own temperature decreases from 650 °C to 279 °C. The 900 °C CO<sub>2</sub> stream is used to heat the feedwater of 145.5t/h from 278.6 °C to 566 °C, while its own temperature decreases from 900 °C to 279 °C. The recovered heat from the carbonator is used to heat the feedwater of 366.6t/h from 278.6 °C to 566 °C. The total recovered heat is about 1008.8MW, in which 465.4MW is from the 650 °C clean flue gas, 154.4MW from the 900 °C CO<sub>2</sub> stream and 389MW from the carbonator.

**Case 3: Use the recovered heat to replace the extracted steam from the turbines and part of the boiler heat load**

It can be found that in Case 1, when the extractions from turbines are all replaced by the recovered heat, there is still part of the recoverable heat not being utilized. In addition, when the heat is all used to replace part of the boiler's load in Case 2, it is not desirable for the stable operation of the system. Therefore, in this case, the recovered heat will be utilized to replace the steam extractions from the turbines and part of the boiler heat to make full use of the recoverable heat from the CO<sub>2</sub> capture process.

It is assumed that the load of the CO<sub>2</sub> capture process is linear with the CO<sub>2</sub> emission. The new heat replacement ratio is defined as:

$$X = \frac{Q_1 + Q_2 + Q_3 - Q_4}{Q_1 + Q_2 + Q_3 + Q_0} \quad (21)$$

Where,  $Q_4$  is the heat used to replace the extractions from turbines.

In Case 3, the condensed water from Point A is divided into eight streams respectively. Approximately, 25% of the boiler heat load is assumed by the recovered heat from the carbonator and CO<sub>2</sub> stream. The recovered heat from the carbonator can replace part of the

boiler load, and the 900 °C CO<sub>2</sub> stream is used to replace the other part of the boiler load. The 650 °C clean flue gas is used to produce steam with the same parameters as the eight extractions ((1)-(8)) in Figure 2. The total recovered heat is about 905.8MW, in which 498.3MW is from the 650 °C clean flue gas, 115.8MW from the 900 °C CO<sub>2</sub> stream and 291.8MW from the carbonator. The output power increase due to less steam extractions is about 148.8MW.

#### **Case 4: Use the recovered heat to build a heat recovery steam generator and produce steam to drive new turbines**

In Case 4, a new steam driven process is built to generate electricity based on the amount of the recoverable heat. It has been calculated that the heat recovered from the CO<sub>2</sub> capture process is able to heat the feedwater of 1255t/h from 278.6 °C to 566 °C, which is enough for a 400MW steam turbine cycle.

In this case, the 650 °C clean flue gas is used to heat the feedwater of 726.5t/h from 278.6 °C to 566 °C, while its own temperature decreases from 650 °C to 279 °C. In addition, the recovered heat from the carbonator is used to heat the feedwater of 528.5t/h from 278.6 °C to 566 °C. The 900 °C CO<sub>2</sub> stream is used to reheat the steam from the high-pressure turbine. The total recovered heat is about 906.6MW, in which 438.6MW is from the 650 °C clean flue gas, 149MW from the 900 °C CO<sub>2</sub> stream and 319MW from the carbonator.

#### **Case 5: Use part of the heat to build a heat recovery steam generator and produce steam to drive new turbines, the other part to replace the extracted steam in the new steam cycle**

In Case 5, a new steam driven process is built to generate electricity with all the extractions from the turbines replaced by recovered heat. It has been calculated that the heat recovered from the CO<sub>2</sub> capture process is able to heat the feedwater of 1025t/h from 248.8 °C to 537 °C, which is enough for a 300MW turbine cycle, while all the extractions from the turbines are replaced by steam generated via recovered heat.

In this case, the 650 °C clean flue gas of 204kg/s is used to provide the desired steam to replace the extractions from the turbines and the 650 °C clean flue gas of 336.1kg/s is used to heat the feedwater of 502.7t/h from 248.8 °C to 537 °C. Additionally, and the recovered heat from the carbonator is used to heat the feedwater of 522.3t/h from 248.8 °C to 537 °C. The 900 °C CO<sub>2</sub> stream is used to reheat the steam from the high-pressure turbine. The total recovered heat is about 777.5MW, in which 303.5MW is from the 650 °C clean flue gas, 158.8MW from the 900 °C CO<sub>2</sub> stream and 315.3MW from the carbonator.

### **4.3 Thermal performance evaluation**

Power output, thermal efficiency, heat consumption rate and coal consumption rate are used as the four indicators in this evaluation, as follows:

#### 1. Power output

The turbine can be divided into different stages, and the real power delivered excluding losses is:

$$W_{out} = F_{in} \times e_{stage} \times \Delta H = F_{in} \times e_{stage} \times (H_{in} - H_{out}) \quad (22)$$

Where,  $H_{in}$  is the inlet steam enthalpy;  $H_{out}$  is the outlet stream enthalpy;  $W_{out}$  is the outlet power per unit; and  $\Delta H$  is the enthalpy change in the stages.

## 2. Thermal efficiency

Thermal efficiency is defined as the ratio of net power generated to the total lower heating value of the consumed fuel.

## 3. Heat consumption rate

The heat consumption rate is defined as the heat needed to generate one kWh of electricity. In all the above cases, the heat needed can be calculated based on the heat supplied by the boiler and the heat consumed in the carbonator.

## 4. Coal consumption rate

The coal consumption rate is defined as the quantity of coal consumed to generate one kWh of electricity. There are various kinds of coal with different heat values, for example, the lower heating values of the two kinds of coal shown in Table 1&2 are different. In order to analyze and compare the total amount of coal consumed, the standard coal needs to be defined. It is defined here that the lower heating value of the standard coal is 29306kJ/kg (7000 kcal/kg) (Tian, 2001; You and Xu, 2008). This definition is especially useful when calculating the cost of the different fuel later in the paper. In all the above cases, the coal consumed should take into account the coal consumption for both the boiler and the calciner.

The conditions of CO<sub>2</sub> streams after heat recovered can be seen from Figures 3. to 7. In all the cases, CO<sub>2</sub> streams will be compressed from atmospheric pressure, at which point it exists as a gas, up to a pressure suitable for pipeline transport (110bar), at which point it is in either the liquid or 'dense phase' regions, depending on its temperature. Therefore, CO<sub>2</sub> undergoes a phase transition somewhere between these initial atmospheric pressure and final pressure (110bar). Compression of the CO<sub>2</sub> to 110 bar will require around 0.4 GJ/tCO<sub>2</sub> (IEA GHG, 2004; IPCC, 2005). In this study, the energy consumption for CO<sub>2</sub> is assumed to be the same for the cases, about 92 MW.

In Table 3, the reference plant which vents its CO<sub>2</sub> to the atmosphere refers to the 600MW power plant without any CO<sub>2</sub> capture process and the reference plant with CO<sub>2</sub> capture but no heat recovery refers to the power plant with a CO<sub>2</sub> capture process but without recovering heat from the CO<sub>2</sub> capture process. The gross power output refers to the power output without deducting the power consumed by the O<sub>2</sub> separation and CO<sub>2</sub> compression processes; while the net power output is the power output excluding the power consumed by the O<sub>2</sub> separation and CO<sub>2</sub> compression processes. The coal consumption rate results in Table 3 are based on the standard coal's heating value.

It can be seen from Table 3 that Case 4 has the most power output, the highest efficiency, lowest heat consumption and lowest total coal consumption rate. Case 2 has the lowest power output. If cases 1 to 5 are compared with the reference plant with CO<sub>2</sub> capture but no heat recovery, their efficiencies increase by 30.4%, 75.8%, 80.4%, 89.7% and 87.1% respectively; while their heat consumptions and their total coal consumptions reduce by 30.4%, 75.8%, 80.4%, 89.7% and 87.1% respectively. The same trend is because there are fixed links among the efficiency, the heat consumption and total coal consumption rate. However, the efficiencies of cases 1 to 5 are much lower than that of the reference plant which vents its CO<sub>2</sub> to the atmosphere due to the large amount of energy consumed by the oxygen separation and CO<sub>2</sub> compression processes. Even for the highest efficiency case among the five cases (Case 4), its efficiency is still 3.8% lower than that of the reference plant which

vents its CO<sub>2</sub> to the atmosphere. The net power outputs of cases 4 and 5 are higher than that of the reference plant which vents its CO<sub>2</sub> to the atmosphere.

	Reference plant which vents its CO <sub>2</sub> to the atmosphere	Reference plant with CO <sub>2</sub> capture but no heat recovery	Case 1	Case 2	Case 3	Case 4	Case 5
Gross power output (MW)	600	600	748.8	600	748.8	1000	989.3
Net power output	600	446	594.8	446	594.8	846	835.3
Efficiency (%)	40.6	19.4	25.3	34.1	35	36.8	36.3
Heat consumption rate (MJ/kWh)	8.87	18.56	14.21	10.56	10.28	9.78	9.91
Total coal consumption rate (g/kWh)	303	634	485	361	351	334	339

Table 3. Comparison of the thermal evaluation results from different cases

#### 4.4 Conclusions

Five cases for the possible integration of a power plant with CaO-based CO<sub>2</sub> capture process have been considered in this paper. Several main conclusions can be drawn as follows:

1. If cases 1 to 5 are compared with the reference plant with CO<sub>2</sub> capture but no heat recovery, their efficiencies increase by 30.4%, 75.8%, 80.4%, 89.7% and 87.1% respectively; their heat consumptions and total coal consumptions share the same trend with the efficiencies, because of the fixed links between the efficiency, the heat consumption and total coal consumption rate.
2. Case 4 has the most net power output of 846MW, the highest efficiency of 36.8%.
3. Case 2 recovers the most total heat, about 1008.8MW but it has the lowest net power output of 446MW and the second lowest of efficiency of 34.1%. This implies that the integrated system's performance depends both on the amount of heat recovery and the type of heat utilization.
4. Case 2 recovers the most heat from the carbonator, while Case 1 recovers the least heat from the carbonator. Case 3 recovers the most heat from the clean flue gas, while Case 5 recovers the least heat from the clean flue gas. Case 5 recovers the most heat from the CO<sub>2</sub> stream, while Case 3 recovers the least heat from the CO<sub>2</sub> stream.

Although the cases in this paper might not be the optimal one in their own category, theoretical factors have been considered in the design of the heat recovery in each case. The optimization of the integration will be included in further study.

#### 5. Conclusions and future work

In this chapter, firstly, CO<sub>2</sub> capture processes based on CaO has been analyzed from the microscopic angle. Secondly, the integrations of CO<sub>2</sub> capture processes based on CaO with

power plants have been discussed. Five cases for the possible integration of a 600MW power plant with CaO-based CO<sub>2</sub> capture process have been discussed. When the system is configured so that recovered heat is used to replace part of the boiler heat load (Case 2), modeling shows that this is the system recovering the most heat of 1008.8MW but also results in the system with the lowest net power output of 446MW and the second lowest of efficiency of 34.1%.

In future work, a dynamic model of the carbonator will be built to analyze the effect of reaction time. The analysis of the integrations has been performed only on a 600MW power plant under base load in this chapter. Power plants with different capacities, for example, 300MW and 1000MW will be taken as the base cases in future research. In addition, the integrations under different working conditions of power plants will be studied as well.

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