The cost structure of a postcombustion CO_2 capture system using CaO

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ABSTRACT

This paper presents the basic economics of an emerging concept for CO_2 capture from flue gases in power plants. The complete system includes three key cost components: a full combustion power plant, a second power plant working as an oxyfired fluidized bed calciner and a fluidized bed carbonator interconnected with the calciner and capturing CO_2 from the combustion power plant. The simplicity in the economic analysis is possible because the key cost data for the two major first components is well established in the open literature. It is shown that there is clear scope for a breakthrough in capture cost

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to around 15 \$/t of CO_2 avoided with this system. This is mainly because the capture system is generating additional power (from the additional coal fed to the calciner) and because the avoided CO_2 comes from the capture of the CO_2 generated by the coal fed to the calciner and the CO_2 captured (as $CaCO_3$) from the flue gases of the existing power plant, that is also released in the calciner.

KEY WORDS: CO_2 capture cost, carbonation, calcination, regenerable sorbent

INTRODUCTION

CO_2 capture from power plants and permanent storage in suitable geological formations can become a major mitigation option for climate change [1]. Most of the key elements in the capture, transport and storage chain are already available and demonstrated at large scale. Capture cost can, therefore, be estimated with reasonable confidence for options really based on "existing technologies" [1]. However, it is widely recognized that there is scope for large reductions in capture cost, and there is a wide variety of R&D efforts around the world to test new concepts to separate CO_2 (or O_2 or H_2). One or more of these large-scale gas separation processes is always at the core of any CO_2 capture system [1].

To get a fair comparison among emerging technologies for CO_2 capture, it is always interesting to analyze the cost structure in these emerging concepts, identifying the critical points for cost reduction and also the dangers for cost escalation. New CO_2 capture technologies have well-defined benchmarks on efficiency and cost, that are established by the "existing technologies" to capture CO_2 [1]. Preliminary cost analysis of emerging CO_2 capture options can help to identify promising paths for development and conclude that some process routes may not have a chance to be competitive against well-established capture options [2] and should be excluded from further support for research and development.

Among the new concepts for CO_2 capture, it is also important to distinguish between two categories. The first refers to totally new concepts, with no analogous reactors at sufficiently large scale in operation today. These technologies require a full scaling up from basic principles tested at laboratory test. The second category refers to new concepts that rely on the use of new functional materials in reactors commercially established at large scale (*e.g.*, a new solvent for absorption columns or a new solid sorbent or gas carrier in a circulating fluidized bed system). Developers of new technologies tend to report cost estimates overly optimistically [1] and always with uncertainty. But it can be argued that the uncertainties are much lower for the second category of new processes. The process discussed in this work falls into this second category.

The CO_2 capture process discussed below was originally proposed by Shimizu *et al*. [3], and uses CaO as a regenerable sorbent to capture CO_2 from combustion flue gases. Other processes using CaO in combustion systems have been proposed [4] but these fall into the first category described above (they require new reactor configurations) and will not be further discussed here. A schematic of the system proposed by Shimizu *et al*. [3] is presented in Fig. 1. CO_2 is captured from the combustion flue gas of an existing power plant in a circulating fluidized bed carbonator operating between 650-700ºC. The solids leaving the carbonator (with a certain conversion of CaO to $CaCO_3$) are directed to a second fluidized bed where calcination/regeneration takes place. Coal burns in the calciner in an atmosphere of O_2/CO_2 at temperatures over 900ºC to produce the heat necessary to calcine the $CaCO_3$ back to CaO and CO_2. This second fluidized bed calciner operates with oxygen supplied by an air separation unit, that consumes power. The CO_2 captured from the flue gases as $CaCO_3$, and the CO_2 resulting from the oxy-fired combustion of coal in the calciner, is recovered in concentrated form in the calciner gas, suitable for final purification and compression (typically >100 bar), for transport and safe storage in a suitable deep geological formation. The compression step marks the boundary of the capture system for cost estimates [1]. The calciner requires a relevant fraction (40-50%) of the total energy entering the system in order to heat up to the calciner temperatures the incoming gas and solid streams and in order to

provide the heat that drives the endothermic calcination of $CaCO_3$. But this energy leaves the system in mass streams at high temperature (at T>900ºC) or is recovered as carbonation heat in the carbonator (at around 650ºC). Therefore, the large energy input to the calciner comes out of the system as high quality heat, available for a high-efficiency steam cycle [3, 5,7]. This is a distinctive characteristic of this capture system, with respect to any other postcombustion CO_2 capture system: it is possible to generate additional power (fraction $(1-f_p)$ in Fig. 1) from the various high-temperature sources in the capture system. The calciner is in fact a new oxyfired fluidized bed power plant (dotted boundary in Figure 1).

Variants of this carbonation-calcination system are planned to be tested at small pilot scale (10s of kW) in ongoing projects in France, UK, US, Germany, Canada and Spain. We have already demonstrated [5] that capturing CO_2 from flue gases with CaO at atmospheric pressure in a batch fluidized bed of CaO can be done with reasonable gas residence times and bed inventories when compared with those in existing large-scale fluidized bed systems. It has also been shown [4] that the potential for high efficiencies exists for a range of these combustion systems. The purpose of this communication is to justify, with a simple and transparent economic analysis, why to expect a very large reduction in CO_2 capture costs when using these CaO loops in post-combustion applications. Lowering the capture cost is the primary driving force to justify the development of a new capture system.

METHODS AND RESULTS

The cost of CO_2 capture can be estimated in different forms, but the most common are incremental cost of electricity ($/kWh) and the cost of avoiding CO_2 ($/t CO_2 avoided).

$$\Delta COE = COE_{capture} - COE_{ref} \tag{1}$$

$$AC = \frac{COE_{capture} - COE_{ref}}{CO_2 kWh_{ref}^{-1} - CO_2 kWh_{capture}^{-1}} \qquad (2)$$

The first equation is self-explanatory when referred to a unit of product ($/kWh) in both the reference plant and the capture plant. This means that the additional energy requirements for the capture step must be incorporated in the estimation of $COE_{capture}$, not only as the extra cost of fuel, but as an increase in capital cost to be able to deliver the same amount of electricity product (kWh). The avoided costs (AC) are derived from equation (2) when the emission factor (tonnes of CO_2 emitted per kWh of electricity produced in the plants) is known. For a more detailed explanation of these and other measures of capture cost, the reader can consult reference [1].

These cost equations require definition of the reference system without capture. The reference system for the purpose of this work will be a power plant using state-of-the-art components similar to those used in the capture system. For simplicity, and because most measures of capture cost in the literature are referred to 1 kWh of electricity, this is the assumed total power output for the system discussed in this work and for the reference plant. The system of reference for this work could also be an existing power plant with sufficient remaining life span, but to simplify the analysis it is assumed that the reference is a newly built pulverized coal or large-scale circulating fluidized bed (CFB) power plant. For any of these combustion systems, reliable capital cost estimates exist (see for example section 3.7 in reference [1] for a comprehensive review of these costs). Also, a set of financial factors (fuel costs, interest rates, *etc*.) and other technical (operating and maintenance costs) and non-technical factors (capacity factor) have to be defined to estimate COE with a simple equation such as:

$$COE = \text{Fixed Costs} + \text{Variable Costs} + \text{Fuel Costs} = \frac{TCR * FCF + FOM}{CF * 8760} + VOM + \frac{FC}{\eta} \qquad (3)$$

The top section of Table 1 summarizes the parameters chosen for the reference system, largely based on data from reference [1]. To introduce flexibility in the assumptions and allow a quick visualization

of their impact on cost figures, three cases have been defined in Table 1 for both the reference plant and the capture system: optimistic set, best estimate and pessimistic set of cost parameters.

Once the reference plant is defined, a capture system of Fig. 1 can also be defined and referred to the production of 1kWhe of electricity. The system of Fig. 1 has three major components, and the same energy output as the reference system (1 kWhe adopted as reference). These include:

1. An existing power plant delivering a major fraction of the total power output, f_p. This should be the largest single component of the total system. The unit cost of producing power in this system is assumed identical to the reference system. This means that it is implicitly assumed that the scale of the existing power plant and the capture system are sufficiently large to be able to benefit from the same economies of scale that apply to the reference system defined in the upper part of Table 1. Therefore, all the economic factors (in particular the normalized capital cost) adopted for the reference case in Table 1 are still valid for the power plant of Fig. 1 delivering f_p kWhe of electricity. Standard equipment for flue gas clean-up in modern power plants is assumed to be included in this component as is the case in the reference system. This is a conservative assumption, as potential synergies (*i.e.*, cost savings) from integrating these components with the capture system (in particular referring to SO_2 and fly ash removal in the CaO loop) are ignored at this stage.

2. A circulating fluidized bed calciner operating as a circulating fluidized bed combustor burning coal in an atmosphere of O_2/CO_2. The oxygen comes from an air separation unit dimensioned to burn only the coal fed to the calciner. The heat requirement for the calciner determines the fraction of coal that is burned in this unit respect to the total. This heat is used to drive the endothermic calcination reaction of $CaCO_3$ and raise all streams flowing to the calciner to the calcination temperature. The high enthalpy of the calcination reaction and the modest conversion of CaO to $CaCO_3$ that is expected in the carbonator, due to modest sorbent performance [3, 4], tend to increase the heat requirements in the calciner. As

discussed in more detailed process simulation work [3-6], f_p should move between 0.5 and 0.6. This is a critical variable for cost estimates as will be shown below. As f_p becomes low, it is increasingly difficult to pursue a system as outlined in Fig. 1, because the same power output could be obtained with a stand-alone oxy-fired CFBC without the added cost of the carbonator described below. However, the fact that a large fraction of the power output in Fig. 1 comes from the existing (cheaper) power plant, is a critical factor in understanding the low cost figures that appear below. On the other hand, oxy-fired CFBC systems are being pursued by some major manufacturers of CFBC equipment [8], and although economic data are scarce, the data adopted in Table 1 are reasonably consistent (slightly more expensive than the expected cost for similar oxy-fired PF boilers) [1]. These figures include all the components required in the oxy fired power plant (CFB boiler, power island, air separation unit, CO_2 compression, *etc*.).

3. A circulating fluidized bed carbonator. This is a large reactor located between the two subsystems described above. It must be dimensioned to treat the combustion flue gases from the power plant, transform part of the CaO into $CaCO_3$ and deliver this solid stream to the calciner after separation from the flue gas in cyclones. It can be considered mechanically similar to existing CFB combustion chambers. This reactor is expected to operate at 600-700ºC at velocities and solid circulation rates typical of CFBCs (see [3, 4, 5]). Since any equipment for heat recovery from the flue gases or solids in this system, and any gas purification equipment, have been implicitly accounted for in the existing power plant or in the oxy-fired calciner, the cost of this piece of equipment can be considered as a small fraction of the capital cost of the power plant described before.

Finally, any other equipment and components making up the capital cost of the full system can be related to the above main components. In particular, the CO_2 purification and conditioning equipment will include a compressor dimensioned to bring to supercritical conditions all the CO_2 flow leaving the calciner (that is the sum of the CO_2 released from the $CaCO_3$ formed in the carbonator and the CO_2

formed in the oxy-combustion of coal in the calciner plus a small fraction coming from the calcination of the $CaCO_3$ make up flow). Also, to estimate the cost of electricity, the maintenance costs (fixed and variable) must be defined. For simplicity and in the absence of more detailed information, the fixed costs have been assumed to be 4% of the capital cost, and the variable operating and maintenance costs (other than the fuel cost) have been assumed to be $0.007/kWh, identical for all cases. As we can see for the reference set of assumptions in Table 1, the cost of electricity is ~$0.039/kWhe for the reference case, and reaches $0.057/kWhe when moving from the reference case to a capture system using oxy-fired circulating fluidized bed combustion. In both reference cases, there is a state-of-the-art steam cycle, delivering 43% LHV efficiency in the case without capture and 32% LHV efficiency in the oxy-fired case (~5% of the net efficiency drop comes from the compression of CO_2 to supercritical conditions and the CO_2 purification, while the other 6% net is due to the air separation unit). The COEs of both the reference case without capture and the case with oxy-fuel combustion are consistent with data collected in [1] and more specifically for CFBC oxy-fired systems in [8]. Capturing CO_2 under these conditions would generate costs between $16-44/tonne CO_2 avoided, depending on the financial assumptions used in the two extreme (optimistic and pessimistic) cases in Table 1.

In order to estimate the COE of the capture system of Fig. 1from the previous cost figures, we need to estimate the power generation efficiency of the new system, proportional to the share of the two major pieces of equipment in the generation of 1 kWh of electricity. We then discount the compression and purification penalty (assumed to be ~5 points of net efficiency) for the CO_2 captured in the carbonator (the penalty for compression of the CO_2 from the coal in the calciner is already included in the efficiency value of the oxy-fuel system). An additional point net efficiency penalty is added to account for purification. The result is :

$$\eta_{capture} = \eta_{reference} \, f_p + \eta_{oxyfuel}(1-f_p) - 0.05 \, f_p \tag{4}$$

With this equation, an intermediate efficiency is obtained for the proposed system, always higher than the efficiency of the equivalent oxy-fired boiler (that would be represented here with $f_p = 0$) and lower than the reference plant without capture (that would be represented here with $f_p = 1$). Efficiency penalties associated with the makeup flow of sorbent required to maintain a given activity in the capture loop [5] are not considered here, as it is argued that an equivalent energy credit would be obtained from a cement plant using deactivated CaO instead of $CaCO_3$ as cement feedstock [4].

In order to calculate the total capital requirement (TCR) of the new system (per kWh, in sufficiently large-scale systems), we use:

$$TCR_{capture} = TCR_{reference} \; f_p \; (1 + \delta_{carb}) + TCR_{oxyfuel} (1 - f_p) \qquad (5)$$

This is a simple function of the TCR of the two main individual components (the power plant of reference and the oxy-fired CFBC power plant used as calciner), adding the carbonator cost (and the extra compression for the CO_2 from f_p) as an incremental fraction, δ_{carb}, of the TCR of the reference plant. This method to calculate the contribution of the carbonator to the TCR of the capture plant is only reasonable for central values of f_p (see Figure 2). If we maintain all the remaining "non technical" parameters of the cost equation (3) as indicated in Table 1, we obtain COEs for the new system that are again somewhere between the extremes for the reference plant and the reference oxy-fired system without carbonation. The avoided costs are calculated from the new emission factor of each system (see Table 1) and yield a central figure of $15.5/tonne CO_2 avoided for the reference case and values below $10/tonne CO_2 avoided for the optimistic case (see Table 1). Figure 2 represents the sensitivity of the cost of electricity in the new capture system depending on the carbonator extra cost parameter and the fraction of total power generated in the power plant of reference.

DISCUSSION

The reduction in capture cost for the new system with respect to the oxy-fired reference case ($f_p = 0$) arises from the assumption that additional (new) power can be extracted from the capture components. Since the separation of CO_2 is carried out at the very high temperatures, all the energy fed to the calciner in the coal feed exits the system in high temperature streams, which are easy to recover (this includes the endothermic heat for calcination, that is recovered as carbonation heat in the carbonator at about 650ºC). This makes the investment per kWhe much lower than for an equivalent stand-alone oxy-fired system. Another way to look at this critical point (not detailed here because it yields similar quantitative results) is to take an oxy-fired CFBC system as the reference in the analysis and think of the carbonator as an interface between this oxy-fired capture system and an existing power plant. The inclusion of a large fluidized bed carbonator in the system boundary of the oxy-fired system will increase (modesly) the capital cost and the COE from this system. But now, the oxy-fired CFBC is not only avoiding the CO_2 from its own coal combustion feed, but all the CO_2 coming from the flue gases of the neighbouring power plant, that has been captured as $CaCO_3$ in the carbonator. In other words, since the oxy-fired CFBC system is capturing about two times more CO_2 than it generates from the combustion of its own coal feed, the avoided CO_2 is higher thanks to the relatively low extra cost of the carbonator unit. This leads to very low avoided costs of capture respect to the stand alone oxyfired power plant.

The previous analysis is highly encouraging for the development of these carbonation-calcination systems. We are aware that the key point for the inmediate future is to demonstrate the technical viability of the system operating at pilot scale in continuous mode. However, as indicated in the introduction, some key results have been completed already [5], and a large body of knowledge exists in the literature on fluidized bed technologies using limestone-derived solids for other purposes. We are, therefore, optimistic about the next steps and about the speed at which results can be escalated, since

the key components are mechanically similar to existing large scale circulating fluidized bed combustors.

NOTATION

COE	levelized cost of electricity (US$/kWh)
AC	cost of CO_2 avoided by using a CO_2 capture system with respect to a similar system without capture (equation 2)
CO_2kWh^{-1}	CO_2 emission factor of the system (kg CO_2 emitted/kWh)
TCR	total capital requirement to build the power plant (US$/kW)
FCF	fixed charge factor (fraction)
FOM	fixed operating costs (US$/yr/kWh)
CF	capacity factor (fraction) and in a 8760 hours typical year
VOM	variable operating costs (US$/kWh)
FC	unit fuel cost (US$/kWh)
η	net plant efficiency (kWhe/ kWh)
f_p	fraction of power generated in the main power plant (asssumed equal to the fraction of coal burned in the main power plant)
δ_{carb}	incremental carbonator cost (fraction)

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Table 1. Summary of assumptions and cost results. Marked in bold are the input data. All remaining cells can be calculated from equations (1-5)

EXISTING POWER PLANT	Units	Optimistic set	Best estimate	Pessimistic set
Fuel Cost, FC	US$ GJ^{-1}	**1**	**1.5**	**2**
Fuel Cost, FC	US$ kWh^{-1}	0.0036	0.0054	0.0072
Capital Cost, TCR	US$ kWe^{-1}	**1100**	**1300**	**1500**
Net efficiency, η_{ref}	kWhe kWh^{-1}	**0.45**	**0.43**	**0.40**
Fixed fraction cost, FOM	-	0.04	0.04	0.04
Capacity factor, CF	-	**0.95**	**0.80**	**0.65**
Fixed charge factor, FCF	yr^{-1}	**0.075**	**0.100**	**0.150**
Variable cost, VOM	US$ kWhe^{-1}	0.007	0.007	0.007
Emission factor, CO_2kWh	kgCO$_2$ kWhe^{-1}	**760**	795	855
Cost of electricity, COE	US$ kWhe^{-1}	0.025	0.039	0.066

OXIFUEL CFBC	Units	Optimistic set	Reasonable set	Pessimistic set
Capture efficiency, E_{oxi}	-	**0.95**	**0.95**	**0.9**
Capital cost, TCR$_{oxi}$	US$ kWe^{-1}	**2100**	**2200**	**2400**
Penalty compression	-	**0.04**	**0.05**	**0.05**
Penalty ASU	-	**0.06**	**0.06**	**0.07**
Penalty total	-	0.1	0.11	0.12
Net efficiency, $\eta_{oxifuel}$	-	0.35	0.32	0.28
Emission factor, CO_2kWh	kgCO$_2$ kWhe^{-1}	48.9	53.4	122.1
Cost of electricity, COE	US$ kWhe^{-1}	0.037	0.057	0.097
Avoided cost, AC	US$ tCO$_2$$^{-1}$	16.4	23.8	44.2

CARBONATOR WITH OXIFUEL	Units	Optimistic set	Reasonable set	Pessimistic set
Carbonator efficiency, E_{carb}	-	**0.9**	**0.8**	**0.7**
Overall capture efficiency, E	-	0.915	0.86	0.80
Carbonator cost fraction, δ_{carb}	-	**0.10**	**0.20**	**0.30**
Capital cost, $TCR_{capture}$	US\$ kWe^{-1}	1566	1816	2175
Conventional power fraction, fp	-	**0.6**	**0.6**	**0.5**
Penalty compression	-	**0.04**	**0.05**	**0.05**
Net efficiency, $\eta_{capture}$	-	0.386	0.356	0.315
Emission factor, CO_2kWh	kgCO_2 kWhe^{-1}	70.9	134.5	217.1
Cost of electricity, COE	US\$ kWhe^{-1}	0.031	0.049	0.089
Avoided cost, AC	US\$ t$CO_2$$^{-1}$	8.3	15.5	36.6

Figure 1 Scheme of power plant incorporating capture of CO_2 with a carbonation loop.
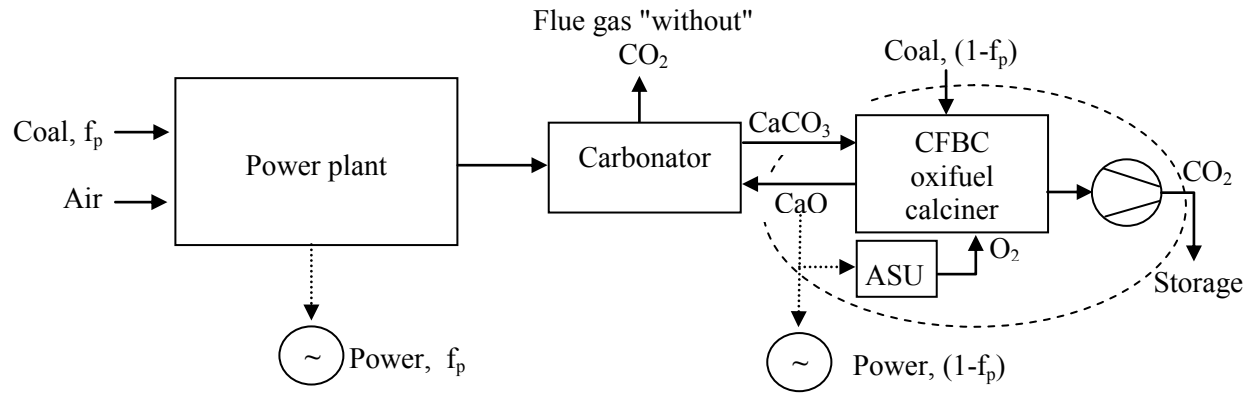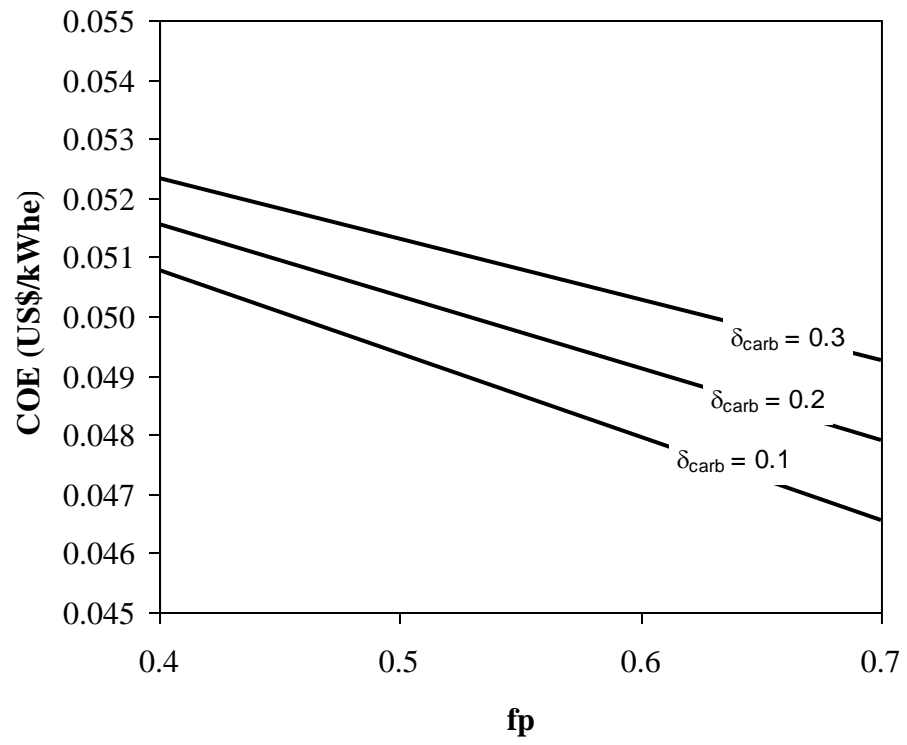
Figure 2: Sensitivity of COE to the fraction of total power output coming from the conventional plant for different carbonator capital cost.

Brief: Postcombustion CO_2 capture using CaO is a low cost emerging technology that doubles the CO2 output of an oxyfired CFBC plant with low efficiency penalty.