BIOMASS COMBUSTION WITH IN-SITU CO₂ CAPTURE WITH CaO. I.: PROCESS DESCRIPTION AND ECONOMICS

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Abstract

A novel concept is proposed to capture CO₂ from biomass, burned in a circulating fluidized bed at a sufficiently low temperature (below 700°C) to allow capture “in situ” by CaO. This step is followed by the calcination of CaCO₃ in an interconnected oxyfuel combustor that delivers a concentrated stream of CO₂ suitable for permanent geological storage. A conceptual design and cost estimation exercise is presented and compared with other similar power plant schemes based on biomass and fossil fuels. The sensitivity analysis of the main cost variables reveals that the concept proposed is competitive against other low carbon systems for lower values of green certificate and higher values of the carbon tax. The process may have long term applications for large scale biomass fired systems and medium term applications in helping large scale power plants with CO₂ capture and storage to meet the target of “zero emissions” by taking advantage of the negative emissions achieved from CO₂ captured from biomass. A second part of this paper demonstrates the concept in a 30 kW circulating fluidized bed test facility.

Introduction

Biomass is regarded as a renewable energy source with zero to low CO₂ emissions if it is produced in a sustainable manner¹. The contribution of biomass and waste to primary energy consumption for power generation in the world is about 1%, in contrast to 81% from fossil fuels². It has been estimated that the contribution of biomass may need to increase to about 5% as part of what is needed to prevent the CO₂ atmospheric concentration from exceeding 450 ppm².

CO₂ capture and storage (CCS) from large stationary sources like power plants is already regarded as a major option for the mitigation of climate change³. It is an acknowledged fact that⁴ the integration of CCS with biomass energy systems leads to CO₂ negative emissions in the generation of electricity or other energy products. However, an assessment of the impact of negative emission energy technologies upon climate change mitigation has only recently been carried out^{5,6}.

Biomass has important physical and chemical similarities with coal, and therefore, the conversion technologies and the final energy products are conceptually similar⁷. However, the use of biomass in the electric sector remains modest, despite the economic incentives on offer, for several reasons. First, there are well known logistical limitations to the use of biomass in large scale power generation systems (see for example Dai et al⁸, and Demirbas et al.⁹). Secondly, there are some fundamental differences in fuel behaviour and ash chemistry that make it quite difficult to switch from one fuel to another in existing combustors in large scale power plants. In large scale coal power plants (> 250 MWe), only

a modest amount of investment is needed to turn existing plants into a co-firing plant^{8, 10} as long as the thermal contribution of biomass to the plant is less than 10%. There are three basic types of co-firing configurations¹⁰: direct co-firing, where biomass and coal are burnt together in the same boiler; parallel co-firing where biomass and coal are burnt in separate boilers; and indirect co-firing where biomass is gasified or combusted separately and the gas produced is injected and burned inside the coal boiler. Direct co-firing is the lowest cost option, but limited by the differences between biomass and coal combustion chemistry¹⁰. Parallel co-firing is limited by the capacity of existing downstream equipment such as that of steam generator¹⁰. Indirect co-firing options are commercially available, but are considered to be too expensive at present¹⁰. However, this last option would be the most suitable option for high ratios of biomass/coal^{8, 10}.

When considering the application of CCS to biomass-based power generation, all the main capture routes (postcombustion, oxyfuel combustion and precombustion) are in principle valid for biomass firing¹¹. Since the postcombustion capture of CO₂ with amine-based solvents is a downstream technology, it is irrelevant whether the source of CO₂ is biomass or coal. The oxycombustion of biomass in pulverized and fluidized bed boilers is also being investigated in co-combustion biomass firing¹². As for pre-combustion, several groups are carrying out research into alternative integration schemes of biomass gasification and CCS^{13, 14}, biomass IGCC and CCS⁷, or co-gasification (through IGCC) and CCS¹⁵. Particularly relevant to the subject of the present study is the Absorption Enhanced Reforming process^{13, 16-18} for hydrogen production from biomass gasification, using CaO as a CO₂ acceptor. As in previous processes involving coal (the Acceptor Process)¹⁹ the CaO particles act as CO₂ absorbent, capturing CO₂ in the gasifier to form CaCO₃, and shifting all

the reforming and water gas shift equilibrium towards H_2 production. The $CaCO_3$ formed in the gasifier is then circulated to a calciner where CaO is regenerated by burning the residual char with air. This gasification process was not specifically designed for CCS applications, because the calciner is air-fired. Recently however, renewed interest has emerged in these technologies which now incorporate regeneration steps that allow oxyfuel combustion to take place in the calciner so that a concentrated stream of CO_2 ²⁰⁻²² can be obtained.

Also relevant to this study is a range of emerging postcombustion CO_2 capture processes that make use of similar calcium loops for CO_2 capture²³⁻²⁵. These processes have certain points in common with other processes that use CaO in relation to sorbent performance and reactor design, as recently reviewed in Blamey et al.²⁶. They are also based on interconnected fluidized bed systems, where a solid flow of CaO is continuously fed into a CO_2 carbonator reactor blown by the flue gases of the power plant and operating at temperatures of around 650°C. The solid stream that leaves the carbonator is directed to an oxyfired calciner (at temperatures over 900°C) where $CaCO_3$ is decomposed back to CaO and CO_2 ready for purification, compression and permanent storage. The carbonator reactor has been shown to work in continuous operation at suitable operating conditions²⁷⁻²⁹. The oxy-fired calciner is a similar reactor to those being developed by the major manufacturers of circulating fluidized bed (CFB) technology^{12, 30, 31} for oxy-fired circulating fluidized bed power plants.

On the basis of the background technologies reviewed so far, this paper discusses a new process for integrating CCS and biomass combustion in power generation. The process uses CaO as a regenerable solid sorbent able to capture CO_2 “in situ” during the combustion of

biomass in a circulating fluidized bed at moderate temperatures. A subsequent step of oxy-fired calcination is able to regenerate the CaCO_3 formed in the first fluidized bed combustion reactor. The concept has been briefly introduced in previous works^{32, 33} but only recently has the first experimental proof of this concept been demonstrated³⁴ in a small pilot plant. In this paper we discuss several scenarios where the concept can be put into practice, including the key operating and design variables. We also consider the cost structure of the process, using as a reference other more mature CO_2 capture processes that use similar equipment. In the second part of this paper, we report on new experimental results obtained in a 30 kW pilot test facility in continuous mode and propose a simple model for interpreting these results on the basis of which valuable design information can be gained for scaling-up purposes.

Process description

The basic scheme of the process which is shown in Figure 1 consists of two interconnected circulating fluidized bed reactors. The combustion of biomass and simultaneous CO_2 capture in the presence of CaO are carried out in the combustor-carbonator reactor. The CaCO_3 formed is regenerated in the combustor-calciner to yield CaO and a rich stream of CO_2 by means of oxy-fuel combustion. The process makes use of the following global reactions:

The basic scheme in Figure 1 may be a stand-alone system (this is perhaps the longest-term option, feasible only for large scale biomass plants) or it may be integrated in different ways with larger fossil-based power generation systems in order to make full use of possible synergies. One option would be to feed the hot flue gas from the combustor-carbonator into the heat recovery section of the existing coal combustion boiler. An added level of integration would be afforded if the system were able to share the oxygen supplied to the combustor-calciner from other larger scale oxyfired coal combustion systems situated nearby. Even more useful would be the availability of a flow of CaO from a larger coal-based oxyfired CFB calciner from other large scale CCS systems that use CaO as regenerable sorbent (see reviews from Blamey et al.²⁶ or Anthony³⁵).

Conceptually, the options mentioned above are able to compete with processes of direct or parallel co-firing of biomass with postcombustion capture of CO₂. However, like the co-firing systems without CCS, these options also have their drawbacks which, for want of space, will not be discussed any further here. However the fundamental operating conditions and the mass and energy flows that are represented in Figure 1 between the two reactors are similar for all options outlined above. These common conditions will be discussed in some detail in the following paragraphs, in order to justify the key mass and energy balances compiled in Table 1 which will serve as a reference set of conditions for the subsequent economic analysis.

The combustor-carbonator illustrated in Figure 1 operates in a combustion environment at atmospheric pressure. The operating temperature in the reactor can be selected on the basis

of the equilibrium of CO₂ over CaO and on whether the simultaneous combustion of the biomass is possible at the temperature chosen. Figure 2 represents the maximum capture efficiency allowed by the CO₂/CaO equilibrium as a function of the combustor-carbonator temperature at different excess air ratios for a typical biomass combustion in air (see composition of biomass in column 1 of Table 1). For all the partial pressures in atm, the maximum CO₂ capture efficiency allowed by the equilibrium ($E_{\text{CO}_2,\text{eq}}$) is:

$$E_{\text{CO}_2,\text{eq}} = 1 - \frac{P_{\text{CO}_2,\text{eq}}(1 - P_{\text{CO}_2})}{P_{\text{CO}_2}(1 - P_{\text{CO}_2,\text{eq}})} \quad (3)$$

where P_{CO_2} is the partial pressure of CO₂ in the flue gas resulting from biomass combustion with a certain excess of air and $P_{\text{CO}_2,\text{eq}}$ is the equilibrium CO₂ partial pressure over CaO estimated by using the equation of Baker³⁶:

$$P_{\text{CO}_2,\text{eq}} \text{ (atm)} = \exp \left(7.079 - \frac{8308}{T(\text{K})} \right) \quad (4)$$

Higher excess air ratios and temperatures tend to lower the capture efficiency allowed by the equilibrium. Our choice of combustor-carbonator temperature is 700°C (with an excess air of 10%). This represents a compromise between a sufficiently high CO₂ capture efficiency (over 80%) and a very high combustion efficiency (assumed to be 100% in all the mass and heat balances that follow).

The temperature of the combustor-calciner reactor chosen was 950 °C, which should allow the fast calcination of the CaCO_3 even in rich atmospheres of CO_2 . In order to carry out the calcination in oxy-fuel conditions, we assumed an oxygen excess ratio of 5% and an inlet oxygen concentration of 25 % vol., which are normal conditions in oxyfired systems when wet CO_2 recycling is used³¹. The heat balance in the calciner has been solved assuming that the gas stream that enters the calciner (stream number 6, which is the CO_2 recycled plus the pure O_2 added) has a temperature of 140 °C¹².

To estimate the specific solid circulation flows between reactors (stream number 4 and 10 in Figure 1) it is necessary to assume a reasonable level of conversion of CaO to CaCO_3 (X_{carb}) in the combustor-carbonator. We have chosen a modest value of $X_{\text{carb}} = 0.1$ and a maximum carbonation conversion or carrying capacity of the CaO of $X_{\text{ave}} = 0.15$. This leaves a remaining active fraction of CaO particles in the combustor-carbonator of 0.05 ($X_{\text{active}} = X_{\text{ave}} - X_{\text{carb}}$). These values of conversion are attainable with natural limestones of low residual activity³⁷, using reasonable make-up flow of fresh limestone³⁸. The ability of the combustor-carbonator to operate with fractions of active CaO of around 0.05 is demonstrated in the second part of this paper³⁹.

One of the key design parameters in this and other calcium looping systems is the heat requirement in the calciner, since that will determine the relative size between the two combustors in Figure 1 and the size of the air separation unit required to supply oxygen to the oxyfired CFBC calciner. For the operating conditions outlined above a heat balance applied to the combustor-calciner would show that 54% of the total heat input to the system must be supplied to the combustor-calciner. Taking into account the low heating value and

carbon contents of the fuels in Table 1, this means that 43.0% of the carbon fed into the system must come from the biomass. The large heat requirements of the calciner are a consequence of the high capture efficiencies in the combustor-carbonator and low sorbent utilization (a large amount of heat is required to heat the non reacted solids from the carbonator up to the temperature of the calciner). It is important to emphasize here that the large heat requirements in the calciner do not entail any significant efficiency penalty for the system (other than the one related to the air separation unit which supplies oxygen to the calciner). This is because the heat released in the calciner can be recovered (in the steam cycle once it has been properly integrated into the system of Figure 1) from the sensible heat of the gas and solids abandoning the calciner and from the carbonation heat released in the combustor-carbonator. The detailed relationship between these variables has been studied in previous works on post-combustion Ca-looping systems⁴⁰⁻⁴⁴.

Finally, it is necessary to define the scale of the reference case in Figure 1 by defining the flow rates of the fuel entering the combustor-carbonator and calciner. Currently, the state of the art of the largest standalone biomass fired combustion power plant is based on circulating fluidized bed technology at a thermal scale between 100 to 250 MWe⁴⁵. Therefore, the mass and heat balance of the calcium looping process in Figure 1 has been calculated assuming a power output from biomass of 250 MWe. This translates into a thermal power input into the combustor-carbonator of 595 MW assuming a typical net power generation efficiency for large scale biomass fired systems of 42%⁴⁵. Coal is assumed to be the fuel in the oxyfired CFBC calciner. The solution of the mass and heat balances in the conditions of Table 1 indicate a heat requirement in the calciner of 705 MWt (0.54 of the total energy input to the system) to calcine the CaCO_3 coming from the

combustor-carbonator, plus the fresh limestone added (stream number 7) and to heat up all incoming streams to the temperature of the calciner⁴⁰.

The other operating variables (gas velocities, solid circulation rates, particle size properties etc) related to the circulating systems in Figure 1 can be considered as relatively standard in commercial circulating fluidized bed combustors without CO₂ capture. This similarity between conventional systems and the one presented here represents an important advantage with respect to alternative emerging CO₂ capture systems. The mechanical and thermal requirements of the key reactors are very similar and well demonstrated in industrial practice including biomass CFB fired systems. Now that the previous set of reference conditions has been established and justified it is possible to analyse the cost structure of the system, taking advantage of its similarity to the commercial components of CFBC units or their emerging CCS oxyfired counterparts.

Cost analysis

The purpose of this section is to explore, by means of a simple and transparent economic analysis, a range of conditions in which the process proposed in the previous section can be economically competitive in producing carbon free electricity. The similarities of the proposed system with existing and emerging CFB power systems allows an economic analysis based on the same methodology as in previous works on Ca-looping processes⁴⁶,⁴⁷. The main task in this section is to provide reasonable estimations of the cost of electricity (COE, \$/kWh) and the cost of CO₂ avoided (AC, \$/t CO₂ avoided) of the low carbon emission technologies chosen for comparison. All cost figures are given as per unit of energy produced and refer to systems of comparable scale (total energy output). Since

biomass based power generation systems are of limited scale, our coal based reference system, against which all the avoided emissions calculated are compared, is a coal based CFBC generating the same electricity output (250MWe) as the system proposed here. The key cost figures for this reference system are listed in the first column of Table 2 (Coal Ref). Also listed in Table 2 are the data corresponding to a second system to be used for comparison: an oxyfired coal based CFBC system (Coal CCS), a third system with a biomass fired CFBC (no capture, indicated as Bio) and a fourth system which is a biomass-oxyfired CFBC (Bio-CCS in Table 2). The key economic figures for the system discussed in this work are displayed in the fifth column (Bio Ca-loop).

An additional set of cost variables (fuel costs, interest rates, *etc.*) and other technical (operating and maintenance costs) and non-technical factors (capacity factor) must be defined in order to be able to estimate COE with the following simple equation:

$$\text{COE} = \frac{\text{TCR} * \text{FCF} + \text{FOM}}{\text{CF} * 8760} + \text{VOM} + \frac{\text{FC}}{\eta} \quad (5)$$

where COE is the cost of electricity, €/MWe, TCR is the total capital requirements, €/kWe, FCF is the fixed charge factor, FOM is the fixed operating costs, (% of capital cost), CF is the capacity factor (%), VOM is the variable operating & maintaining cost, €/kWe, FC is the fuel cost, €/kWh, and η is the net efficiency (kWe/kWh).

For the reference case without capture (Coal Ref) and the biomass case without capture (Bio) data available in public literature can be used to estimate COE. In this work, we have chosen the values reported in Table 2. These are consistent with recent sources of CFBC

power plant cost data (cost calculator recently published by the European Commission⁴⁸) and a recent study carried out by Foster Wheeler for the IEA Greenhouse Gas R&D Programme⁴⁵. The total capital requirements TCR for the biomass CFBC power plant (column “Bio”) are assumed to be the same as those of the air-fired coal CFBC (Coal Ref in Table2), except in the case of fuel cost and efficiency (0.45 vs 0.42).

The cost data for the coal CFB power plant with capture (Coal CCS in Table 2), corresponding to the same electricity output as the reference, are obtained assuming oxy-combustion CFB technology. These Oxy-fired CFBC systems are being pursued by some major manufacturers of CFBC equipment^{12,31} and although economic data are scarce, the data presented in Table 2 are reasonably consistent with those found in the literature on oxy-fired circulating fluidized bed boilers (slightly more expensive than similar oxy-fired pulverized fuel boilers). These figures include all the components required in an oxy fired power plant (CFB boiler, power island, air separation unit, CO₂ compression, *etc.*). As in the case of no capture, the cost data for a biomass oxyfired CFBC power plant (column “Bio-CCS” in Table 2) are assumed to be the same as those of the Coal CCS, except in the case of fuel cost and efficiency (0.33 vs 0.30).

Finally, the cost of the capture system in Figure 1 (Bio Ca-loop in Table 2) can be calculated by following the same strategy as that adopted in a previous work for postcombustion CO₂ capture systems⁴⁶ bearing in mind that the system in Fig. 1 has two major components:

1. The combustor-carbonator can be considered as a conventional biomass CFBC boiler when it comes to cost estimation because there are no relevant differences in operating conditions except in the operation at slightly lower temperatures. All the materials used in its construction should be identical to those of a conventional boiler with the same biomass input. There will only be a marginal increase in the heat exchange areas necessary to extract the heat from the combustion chamber, as the usual temperatures in commercial boilers are about 50-150°C higher than the temperature chosen for the combustor-carbonator (700 °C).
2. The combustor-calciner is a circulating fluidized bed combustor that burns a solid fuel in an atmosphere of O₂/CO₂. The oxygen comes from an air separation unit dimensioned to burn only the fuel fed into the calciner. The heat requirement for the calciner (set at 54% of the total, as shown in Figure 1 and Table 1) determines the fraction of fuel that is burned in this unit in proportion to the total.

Any other equipment included in the full capture system (Bio Ca-loop) that increases the capital cost of the overall system can be related to the above main components. In particular, the CO₂ purification and conditioning equipment will include a compressor designed to bring to supercritical conditions all of the CO₂ flow that leaves the calciner (that is the sum of the CO₂ released from the combustor-carbonator, the CO₂ resulting from the oxy-combustion of coal in the calciner and the small fraction of CO₂ produced by the calcination of the CaCO₃ make up flow). Treatment of exhaust gases are not considered within this work. As shown in Paper II, the flue gas of the combustor carbonator may include unconverted gaseous components. This may lead to a high demand of energy after treatment (even post combustion chamber or catalysts) and therefore higher investment costs that are not included in the capital cost of this system. In addition, to estimate the cost

of electricity, it is necessary to define the maintenance costs (fixed and variable). For the sake of simplicity the fixed costs have been assumed to be 3.7% of the capital cost⁴⁸ as in similar CFBC power systems, and the variable operating and maintenance costs (other than the cost of fuel) have been assumed to be \$0.01 €/kWh, in all cases⁴⁸ for the same reason.

In the two cases without CCS, there is a state-of-the-art steam cycle, that delivers 45.0% LHV efficiency in the coal CFB reference power plant (Coal Ref)⁴⁸ and 41.7% LHV efficiency⁴⁵ in the biomass-fired plant without CO₂ capture (Bio). In the oxy-fired case (Coal Oxy), efficiency drops to 33% LHV (around 5% of the net efficiency drop resulting from the compression of CO₂ to supercritical conditions and CO₂ purification, while the other 7% net is due to the air separation unit, with respect to the Coal Ref system). An intermediate efficiency penalty of 35% is obtained for the proposed system (Bio Ca-loop). This is lower than the efficiency of the equivalent reference biomass CFB plant without capture (Bio in Table 2) represented by the system of Figure 1 with only one combustor but higher than that of the stand-alone oxy-fired boiler (Coal CCS) represented by Figure 1 with only one combustor-calciner. Efficiency penalties associated with the calcination of the makeup flow (solid stream 7 in Figure 1) of limestone required to maintain a given activity in the capture loop are considered zero in this work, as it can be argued that an equivalent energy credit could be obtained from a cement plant using deactivated CaO instead of CaCO₃ as cement feedstock⁴⁹.

With all the previous assumptions in place, the COE of the resulting systems has been calculated without any incentive (“NI” cells in Table 2).

The avoided cost can now be estimated for the low carbon (“low C”) power generation systems, with respect to the reference systems in the first column, $\text{COE}_{\text{CoalRef}}$ and $\text{CO}_{2\text{CoalRef}}$ kWh^{-1} , as follows:

$$\text{AC}_{\text{lowC}} = \frac{\text{COE}_{\text{lowC}} - \text{COE}_{\text{CoalRef}}}{\text{CO}_{2\text{CoalRef}} \text{kWh}^{-1} - \text{CO}_{2\text{lowC}} \text{kWh}^{-1}} \quad (6)$$

This equation is self-explanatory if the “low C” system is the Coal CCS. Then the specific emissions in the reference plant ($\text{CO}_{2\text{CoalRef}} \text{kWh}^{-1}$) and the capture plant ($\text{CO}_{2\text{CoalCCS}} \text{kWh}^{-1}$) can be defined and all the additional energy requirements for the capture process can be incorporated in the estimation of COE_{lowC} . The same equation can be used to compare the reference case against biomass-fired systems without capture by changing the notation “lowC” to “Bio” and taking into account that the CO_2 emitted from the combustion of the biomass ($\text{CO}_{2\text{bio}} \text{kWh}^{-1}$) is considered “neutral” and therefore has to be taken out of the denominator:

$$\text{AC}_{\text{Bio}} = \frac{\text{COE}_{\text{Bio}} - \text{COE}_{\text{CoalRef}}}{\text{CO}_{2\text{CoalRef}} \text{kWh}^{-1}} \quad (7)$$

When the CO_2 is captured from the combustion of biomass, equation 6 must be adjusted to discount the CO_2 from biomass that is captured as it can be considered as “negative emission”. In other words, when the CCS system (oxycombustion CFBC) is applied to biomass, the specific emission of the low C system ($\text{CO}_{2\text{lowC}} \text{kWh}^{-1}$) for equation 6 is defined as:

$$\text{CO}_{2\text{BioCCS}} \text{kWh}^{-1} = -\text{CO}_{2\text{Bio}} \text{kWh}^{-1} \frac{\eta_{\text{Bio}}}{\eta_{\text{BioCCS}}} E_{\text{oxy}} \quad (8)$$

where the $\text{CO}_{2\text{Bio}} \text{kWh}^{-1}$ is the CO_2 emission factor associated to a biomass combustor, η_{Bio} and η_{BioCCS} are the net efficiencies of the Bio plant and the Bio CCS plant and E_{oxy} is the

capture efficiency assumed for the biomass oxy-combustor. Note that this is a negative value that accounts for all of the CO₂ captured and stored from the biomass.

Finally, in the system analyzed in this work, only a fraction of the total amount of carbon fed into the system comes from the biomass ($f_C=0.43$ as pointed out previously) and it is captured (a $E_{carb}=80\%$ capture is assumed in the combustor-carbonator). There is also a certain amount of CO₂ emitted from the coal used in the calciner that is not captured (a $E_{oxy}=95\%$ capture is assumed for the oxyfired calciner). This translates to a specific emission factor for CO₂ in equation 6 of:

$$CO_{2BioCaLoop} kWh^{-1} = -CO_{2Bio} kWh^{-1} \frac{\eta_{Bio}}{\eta_{BioCaLoop}} f_C E_{carb} + CO_{2CoalRef} kWh^{-1} \frac{\eta_{CoalRef}}{\eta_{BioCaLoop}} (1 - f_C) (1 - E_{oxy}) \quad (9)$$

where $\eta_{BioCaLoop}$ and $\eta_{CoalRef}$ are the net efficiencies of the Bio Ca-loop and the Coal Ref plant, respectively.

Until now the electricity cost and avoided cost of all the systems in Table 2 have been calculated without incentives. In order to analyze the effect of incentives on these systems, two additional cost scenarios are calculated in Table 2, assuming the same incentives for CCS systems and biomass use in power generation as those assumed in a recent IEA-GHG Report⁴⁵. One includes a price for the CO₂ emitted under the ETS (Emission Trading System) market in Europe. The ETS price penalizes CO₂ emissions from fossil fuels. The other cost scenario incorporates a representative ETS price of 14 €/t CO₂ and a Green Certificate (GC) adopted in several locations in Europe, of 50 € of profit per MWh of electricity produced by biomass combustion. For consistency, this value is again adopted from the IEA-GHG Report⁴⁵.

For the central set of conditions adopted in Table 2, the resulting cost under different incentives is compared in Figure 3a. As expected, the scenario with no incentives shows that the cost of electricity from biomass based systems, with or without capture, is substantially higher than the cost of coal based systems mainly due to the much higher cost of the fuel. Particularly high is the cost of electricity from biomass with CCS (Bio CCS column) even compared with coal with CCS. This is because, in addition to the capital increment associated with CCS, even though the energy penalty is assumed to be the same for both CCS systems, in the case of Bio CCS the additional fuel needed to compensate for the energy penalty is much more expensive. The system proposed in this work, Bio Ca-loop, benefits in these conditions from both the inherent lower energy penalties reported in Table 2 and from the use of coal (not biomass) in the calciner. When the ETS price of 14€/tCO₂ is introduced into the equations, the COE_{CoalRef} becomes more expensive, while Coal CCS increases its cost only slightly (due to the small flow of CO₂ not captured that is emitted from the plant with capture). However, in the case of Bio CCS, there is a drop in the cost of electricity because it is assumed that the flow of CO₂ captured from biomass will be able to claim a credit from the ETS scheme (negative emission, as indicated in equation 8). The process presented in our work also benefits from this ETS incentive but to a lesser extent than Bio CCS, because the biomass is assumed to be burned only in the combustor-carbonator whereas the coal is consumed in the combustor-calciner.

When an additional incentive is introduced to encourage a wider use of biomass for electricity production (Green Certificate of 50€/MWh), a large distortion in the electricity prices takes place. As shown in Figure 3a, this particular value of the incentive favours the

use of biomass so much that the lowest electricity cost comes from the biomass fired power plant without capture, “Bio” in Figure 3a.

When the avoided cost of CO₂ (Figure 3b) with no incentives is considered, the lowest avoided cost corresponds to the Coal CCS system, mainly because of the impact of the lower fuel price and energy penalty on the overall cost. For the same reasons, the highest avoided cost corresponds to the biomass power plant without capture. The introduction of CCS to the biomass systems (Bio CCS and Bio Ca-loop) translates into a sharp decrease in avoided cost because of the effects of negative emissions resulting from capture mentioned in the previous paragraph. It is interesting to note that the lower energy penalty and lower specific capital cost of the Bio Ca-loop systems yields a slightly lower avoided cost than the Bio CCS despite the lower amount of CO₂ capture from biomass. If biomass could be used in the combustor-calciner of Figure 1, the avoided cost of the Bio Ca-loop would decrease to 0.3 €/t CO₂. Finally, it should be mentioned that the effects of ETS on avoided cost are straightforward for all systems because there is a drop in avoided cost equal to the ETS credit of 14 €/tCO₂ avoided. If the Green Certificate incentive is included, the benefit is more substantial in the stand-alone biomass plant (named Bio) due to the low cost of electricity derived from this credit which makes the COE of the biomass system (Bio) lower than that of the coal based reference (Coal Ref).

Figure 4 shows the sensitivity of the cost of electricity and avoided cost to the variation of the incentives considered above. As was previously pointed out, the cost of electricity derived from biomass fired systems is favourable when GC is incorporated into the economic assessment. It is to be noted that the lowest cost of electricity among the biomass

systems is obtained for the Bio Ca-loop plant when the revenues from the GC are lower than 20 €/MWh and an ETS of 14 €/t CO₂ is assumed. However, when the GC is given values higher than 40 €/MWh, the Bio system is the most competitive. Once more, the variation in costs of CO₂ avoided (AC) shows that biomass fired systems benefit from the Green Certificate reward. In a scenario with high incentives for the use of biomass, the stand-alone biomass power plant is the most favoured energy system, accounting for net revenues of more than 40 €/ t CO₂ avoided.

Regarding the sensitivity of the COE assessment to the ETS mechanism (assuming a GC of 50 €/MWh) it should be noted that the Bio CCS plant, whose COE is the highest if no ETS incentive is applied, would benefit the most if the ETS price reached values higher than 50 €/t CO₂. In this case, the Coal Ref system would be the worst option since it leads to an increase in the COE in the ETS market. However, the Bio plant does not receive any advantage from this credit since its emissions are considered as neutral. As regards avoided costs, all of the low carbon options show the same trend. The stand alone biomass plant is the most favoured since the difference in COE compared to the reference case (Coal Ref) is higher for high ETS credit values.

Conclusions

The equilibrium of CO₂ over CaO allows an effective “in situ” capture of the CO₂ generated during biomass combustion at around 700°C in a circulating fluidized bed. A comprehensive CO₂ capture process can be designed if it is coupled with a calcination step in which coal is burnt under oxyfired conditions. The similarity of this process to existing or emerging CFB combustors makes it possible to estimate the main costs of the process on

the basis of reasonable assumptions, carbon prices (European Trade System, ETS) and incentives for the use of biomass for power generation (Green Certificates). The lowest cost of electricity with the ETS incentive of 14 €/t corresponds to a coal based power plant without CCS whereas the lowest cost of electricity with a Green Certificate of 50 €/MWh and an ETS of 14 €/t CO₂ corresponds to a stand-alone biomass power plant. As for avoided cost, the stand-alone biomass plant shows the lowest cost when it is rewarded with ETS and the green certificate. The system proposed in this work is competitive, and it is only surpassed by the stand-alone biomass system, when there is both an incentive for electricity production from biomass and a price on carbon. A sensitivity analysis of the main cost variables in relation to these incentives reveals that the proposed concept can be competitive against other biomass systems with lower values of green certificate and higher values of ETS credit.

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Nomenclature

AC	Avoided cost for low carbon systems, €/tCO ₂ avoided
CF	Capacity factor, %
CO ₂ kWh ⁻¹	CO ₂ emission factor, tCO ₂ /kWh
COE	Cost of electricity, €/MWh
E _{carb}	CO ₂ capture efficiency in the combustor-carbonator
E _{CO₂,eq}	Maximum CO ₂ capture efficiency
E _{oxy}	CO ₂ capture efficiency for a biomass oxy-combustor
FC	Fuel cost, €/kWh
fc	Carbon fraction from biomass
FCF	Fixed charge factor
FOM	Fixed operating cost, % of TCR
h	Net efficiency, kWh/kWh
P _{CO₂}	Partial pressure of CO ₂ , atm
P _{CO₂,eq}	Equilibrium CO ₂ partial pressure over CaO, atm
TCR	Total capital requirements, €/kWh
VOM	Variable operating & maintaining costs, €/kWh

Subscripts

lowC	Low carbon system
CoalRef	Coal reference system
Bio	Biomass plant without capture
BioCCS	Biomass plant with CO ₂ capture and storage
Bio CaLoop	Biomass plant with Ca-Loop

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Figure 1. General scheme of the proposed process (composition of the streams for the reference case compiled in Table 1)

Figure 2. Variation of the maximum carbonation efficiency allowed by the equilibrium of CO_2 over CaO with temperature at 0, 10 and 40 % excess air ratios.

Figure 3. Comparison of (a) the cost of electricity and (b) the avoided cost for different scenarios of incentives.

Figure 4. Comparison of the variation of the cost of electricity and the cost per unit of CO₂ avoided for different technologies as a function of the Green Certificate credit (graphs on the left; ETS = 14 €/t CO₂) and the ETS price (graphs on the right; Green Certificate = 50 €/MWh)

Stream	1	2	3	4	5	6	7	8	9	10
T (°C)	20	20	700	700	20	140	20	950	950	950
F (kg/s)	33.6	180.2	172.5	634.1	26.2	268.5	8.4	5.3	127.8	592.7
LHV (MJ/kg)	18				27					
H* (MW)	0.0	0.0	145.9	446.3	0.0	29.7	0.0	5.1	167.5	561.2
Comp. (% wt)										
C	41.4				70.4					
H	5.4				4.9					
S	0.0				0.0					
O	39.9				13.0					
H ₂ O	7.2		10.9		9.0	8.6			10.9	
N	0.7				1.3					
Ash	5.4		0.7	11.6	1.4	0.2		12.3	0.2	12.3
O ₂		26.3	2.2			21.9			0.8	
N ₂		73.7	80.3			0.2			0.3	
CO ₂			5.9			69.1			87.8	
CaCO ₃				14.6			100.0			
CaO				73.8				87.7		87.7

Table 1. Composition and general characteristics of the different streams in Figure 1 for the reference case. (* Sensible heat of the mass stream with respect to the reference at 20°C).

Table 2. Main parameters assumed and comparison of cost estimations for different technologies

		"low C" systems														
		Coal Ref			Coal CCS			Bio			Bio CCS			Bio Ca-loop		
Coal Cost, FC_{coal}	€/kWh	0,010			0,010									0,010		
Biomass Cost, FC_{bio}	€/kWh							0,030			0,030			0,030		
Fraction of power from carbonator-combustor, f_p														0,46		
Fraction of carbon fed as biomass, f_c														0,43		
Capital Cost, TCR	€/kWe	1400			2500			1400			2500			1994		
Fixed Fraction Cost, FOM	(% of CC)	3,70			3,70			3,70			3,70			3,70		
Capacity Factor, CF	%	90			90			88			88			88		
Fixed Charge Factor, FCF		0,10			0,10			0,10			0,10			0,10		
Variable Cost, VOM	€/kWhe	0,01			0,01			0,01			0,01			0,01		
Carbonator Capture Efficiency, E_{carb}														0,80		
Oxy-combustor Capture Efficiency, E_{oxy}					0,95						0,95			0,95		
Overall Capture Efficiency, $E_{capture}$														0,88		
Penalty ASU					0,070						0,070			0,038		
Penalty Purification and Compression					0,050						0,050			0,050		
Penalty Use of Biomass								0,033			0,033			0,015		
Total Penalty					0,120			0,033			0,153			0,103		
Net Efficiency, η	kWhe/kWh	0,45			0,33			0,42			0,30			0,35		
CO ₂ Emission Factor, CO ₂ kWh ⁻¹	t CO ₂ /kWhe	765			52			0			-971			-273		
		NI*	ETS†	GC‡ & ETS	NI	ETS	GC & ETS	NI	ETS	GC & ETS	NI	ETS	GC & ETS	NI	ETS	GC & ETS
Cost of Electricity, COE	€/MWhe	52	62	62	75	75	75	101	101	51	145	132	82	93	89	66
Avoided Cost, AC	€/t CO ₂ avoided				32	18	18	65	51	-14	54	40	11	40	26	4

* NI: No incentives

† ETS = 14 €/t CO₂ (penalty if CO₂ from coal is emitted; revenue if CO₂ from biomass is captured)

‡ GC = 50 €/MWhe of electricity produced by biomass