

Thermal Integration of a Flexible Calcium Looping CO₂ Capture System in an Existing Back-Up Coal Power Plant

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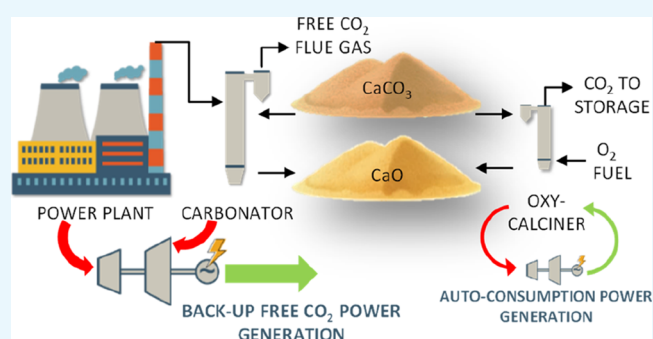


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ABSTRACT: The CO₂ capture from back-up power plants by making use of calcium looping systems combined with large piles of Ca-solids has been studied in this work. A flexible CO₂ capture system based on a concept described in a previous work has been integrated into an existing power plant by including a small oxy-fired calciner (that represents just 8% of the total thermal capacity) to steadily regenerate the sorbent and a carbonator reactor following the back-up power plant operation periods to capture 90% of the CO₂ as CaCO₃ and two large piles of rich CaO and CaCO₃ solids stored at modest temperatures. When the back-up plant enters into operation, the calcined solids are brought into contact with the flue gases in the carbonator reactor; meanwhile, the oxy-calciner operates continuously at a steady state. In order to improve the flexibility of the CO₂ capture system and to minimize the increase of CO₂ capture costs associated with the additional new equipment used only during the brief back-up periods, we propose using the steam cycle of the existing power plant to recover a large fraction of the heat available from the streams leaving the carbonator. This makes it possible to maintain the electrical power output but reducing the thermal input to the power plant by 12% and thus the size of the associated CO₂ capture equipment. To generate the auxiliary power required for the oxy-calciner block, a small steam cycle is designed by integrating the waste heat from the streams leaving this reactor. By solving the mass and heat balances and proposing a feasible thermal integration scheme by using Aspen Hysys, it has been calculated that the CO₂ emitted by long-amortized power plants operated as back-up can be captured with a net efficiency of 28%.



1. INTRODUCTION

Renewable power is expected to be the dominant energy source in future energy networks, with shares of the total electricity demand of between 75 and 80% by 2060.^{1,2} Back-up combustion and/or energy storage systems will be required to maintain the balance between supply and demand in such systems.^{2,3} Existing amortized coal power plants operated as back-up could be favored in such future electricity grids by market mechanisms, as they use a relatively low-cost fuel and can be largely amortized.^{3–6}

However, the decarbonization of these combustion back-up power plants by means of CO₂ capture and storage (CCS) technologies will be needed to accomplish the aggressive CO₂ emission reduction targets established for the next few decades.⁷ CCS systems have a limited flexibility to cope with the fluctuations in the flue gas loads and, what is more, their high investment costs seriously limit their applicability at low capacity factors (CFs).^{8–14} A number of solutions have been proposed in the literature to increase the flexibility of the CO₂ capture systems, most of which involve the storage of different functional materials. The storage of a fraction of the rich solvent has been studied in post-combustion amine-based CO₂ capture systems^{14–16} to reduce the consumption of power in the regeneration and CO₂ compression units (CPU) during

the periods of peak power demand, so that these operations can be postponed to when there is less demand. The storage of O₂ in cryogenic tanks during low power demand periods while operating the air separation unit (ASU) in base mode has been proposed for oxy-combustion systems.^{17–19} Similarly, the storage of H₂ in pre-combustion CO₂ capture systems has been analyzed to decouple the generation of power from H₂ production.^{20,21} However, these approaches are aimed at achieving short-term flexibility (i.e., ranging from minutes to hours) to cover short peaks characterized by a fast response. In the case of seasonal back-up (i.e., operating only a few weeks per year), such CO₂ systems will be penalized due to the required large-capacity storage of costly materials (such as amines) or due to the use of large-scale storage equipment (i.e., for cryogenic O₂ or pressurized H₂).

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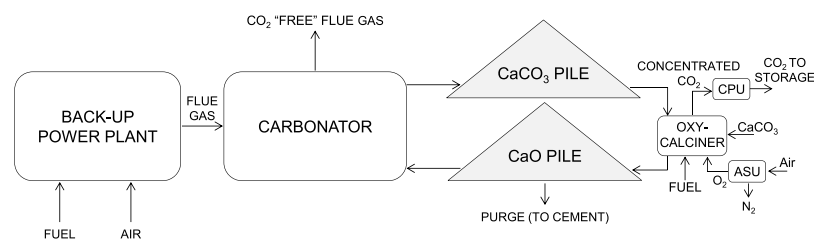


Figure 1. Schematic representation of the highly flexible back-up fossil power plant with CO₂ capture by CaL including CaO/CaCO₃ storage (FlexiCaL).

A CO₂ capture process to improve the flexibility of calcium looping (CaL) technology has been recently proposed in order to take advantage of the low specific cost of the solids required for large-scale storage, that is, CaO and CaCO₃ derived from crushed limestone.²² CaL technology has advanced rapidly in the last decade and it has been demonstrated at TRL6 in several pilot plants.^{23–28} The similarity of its core components, the circulating fluidized bed (CFB) carbonator and calciner reactors, with existing CFB boiler technology has led to several testing and modeling studies,^{29–33} process simulation, and integration^{34–38} as well as cost analyses^{33,39–41} in a wide range of configurations. Despite the progress achieved in the CaL technology, the flexibility of CaL systems against flue gas load changes has received relatively low attention until recently.^{42–44} Some authors have also simulated CaL-based energy storage systems for CCS^{45,46} and solar power applications,⁴⁷ respectively, by making use of the thermal and/or thermochemical energy contained in the streams and/or reactions of the CaL system.

In the present work, we investigate the thermal integration of a particular reference flexible CaL case that could be used to extend the life of existing coal-fired power plants operated as back-up power systems. The main goal for this system is to minimize the cost of the carbon-free back-up power service by reducing the capital investment associated with the CO₂ capture. In addition, this process aims to moderate as much as possible energy penalties and to concentrate them in periods of low power demand. For this purpose, the use of the existing power plant steam cycle to integrate the heat available from the carbonation reaction has been studied in this work as a mean to reduce the cost of the CaL equipment, which is essential for its viability when operating under the low CFs expected in any decarbonized back-up power system.

2. PROCESS DESCRIPTION

The general scheme of the CO₂ capture system analyzed in this work (referred to as “FlexiCaL” from this point on) is presented in Figure 1. This process presents features in common with a conventional CaL configuration, such as the use of two CFB reactors, the carbonator and the calciner, but it also includes additional features to ensure the flexible operation of the CO₂ capture system, as was discussed in our previous work.²²

As shown in Figure 1, one of the main characteristics of the FlexiCaL system is the use of two low-cost piles (or silos) of CaO- and CaCO₃-rich material (referred to as CaO and CaCO₃ piles for simplicity, although other minor components such as sulfates or ash may also be present). These piles could be dimensioned as to allow the complete decoupling of the carbonator and calciner operation modes. Thus, calcined solids from the CaO pile are fed directly into the carbonator to react

with the CO₂ present in the flue gas when the back-up power plant enters into operation. Meanwhile, the carbonated solids leaving this reactor are stored in the CaCO₃ pile and the “free” CO₂ gas stream is released. The FlexiCaL process in Figure 1 aims to minimize the temperature of the piles in order to facilitate the handling and storage of the CaO and CaCO₃ materials. As shown below, different integration schemes can be used to achieve this objective, reducing the temperature of the solids in the piles to below 250 °C.

As a high-temperature solid looping system, typical calcium looping schemes are characterized by its ability to recover waste heat in a boiler, not only from the exothermic carbonation reaction that typically takes place at 650 °C⁴⁸ but also from the integration of the gas and/or solids streams leaving the carbonator and calciner reactors. Unlike most CaL schemes that use boiler-type carbonators, the carbonator reactor in the process depicted in Figure 1 is assumed to be an adiabatic reactor from which no heat is recovered for power production. However, a significant amount of thermal energy can be recovered from the gas and solids streams leaving the carbonator reactor to produce power. Because the process in Figure 1 only targets extremely low CFs, the use of an additional steam cycle linked to the gas and solids streams from the carbonator is ruled out in this work in an attempt to minimize waste of capital. Therefore, a new approach is adopted in this work to recover part of the energy from the carbonation using the steam cycle of the existing back-up power plant. The availability of such thermal capacity from the carbonation allows a certain reduction of the coal thermal input. This leads to a reduction in the flow of CO₂ produced in the power plant and, therefore, in the amount of sorbent needed in the carbonator. Different integration schemes can be adopted to incorporate both the thermal power from the power plant and the carbonation heat available into a single steam cycle. In standard CaL schemes, some of these options were investigated by Yang et al.³⁶ However and in order to minimize the modifications in the power plant steam cycle and, thus, the capital cost, we have considered only the use of the carbonation heat to preheat the water entering the boiler, as discussed in the following section.

3. RESULTS AND DISCUSSION

A reference case is proposed to illustrate the performance of the FlexiCaL system of Figure 1. The power plant, carbonator, and oxy-calciner, as well as the associated steam cycles have been modeled using Aspen Hysys to solve the mass and energy balances in steady-state mode; meanwhile, the nonsteady operations related with the switching on and off of the power plant and carbonator block for the back-up periods has been left outside the scope of this work. Relevant parameters and values for the reference case described below are summarized

in Table 1 and the different major elements in Figure 1 are detailed below.

Table 1. Main Parameters for the Reference Case Studied in This Work

description	unit	value
net electrical power output, $P_{e,net}$	MW _e	350
power plant thermal capacity, $P_{th, power\ plant\ ref}$	MW _{th}	777
net power plant efficiency, $\eta_{power\ plant}$	%	45
CF		0.1
carbonation temperature, T_{carb}	°C	650
oxy-calcination temperature, T_{calc}	°C	910
average CO ₂ carrying capacity, X_{ave}		0.35
CO ₂ capture efficiency, E_{carb}		0.9
calcination efficiency, E_{calc}		1

As general considerations for the analysis in the following sections, it has been assumed that heat losses in the solid piles can be neglected considering their low surface-to-volume ratio, and in gas–solid heat-exchange operations taking place in cyclonic preheaters, the outlet stream temperatures are equalized and the heat exchange efficiency in the steam cycles is of 95%.

As in other CaL systems and as considered for the reference case here presented, an additional benefit of the FlexiCaL process in Figure 1 is its potential synergy with a large-scale CaO consumer such as a cement plant. As in typical CaL, a certain flow of fresh limestone (make-up flow, F_0) is fed into the calciner in order to compensate for the decay in the sorbent's CO₂ carrying capacity and also to purge the inserts from the inventory of solids (ashes and CaSO₄ formed in the oxy-calciner). In addition, the low CFs of the system in Figure 1 allows for very high activity materials in the solid storage piles when employing make-up flows of limestone that satisfy the requirements of a typical cement plant.

3.1. Reference Coal-Fired Power Plant. An existing coal-fired power plant providing a net electric power of 350 MW_e is considered to be the back-up system and the source of the flue gas fed into the carbonator. This power plant employs a supercritical steam cycle operating with live steam at 600 °C and 280 bar. Figure 2 shows a simplified scheme of the steam cycle of the existing power plant. The main operation conditions of the steam cycle have been adopted from the data available in the literature⁴⁹ and assuming that a 5% of the gross power output is consumed by the auxiliaries, resulting in a net power efficiency of 45% (defined as the ratio between the net electrical output and the thermal input to the power plant). As a back-up system, the power plant is assumed to run at full load when it enters into operation, but with a low CF of 0.1 averaged over 1 year. During the operation periods, a coal thermal input of 777 MW_{th} is fed into the power plant, producing a flue gas flow of 10.4 kmol/s at 140 °C with 15.3 % v of CO₂. This power plant is assumed to be equipped with a desulfurization unit and, therefore, no SO₂ is emitted with the flue gas.

3.2. Back-Up CO₂ Capture and Thermal Integration of the Carbonator Block. A CO₂ capture efficiency of 90% has been chosen as the target in the carbonator operating at a temperature of 650 °C. An average CO₂ carrying capacity (X_{ave}) of 0.35 has been assumed. This value has been taken by considering that the flow of fresh limestone to the oxy-calciner reactor (F_0) is continuous (even when the carbonator and hence the power plant are not operating). This results in a flow of CaO particles with a high activity accumulating in the CaO pile that only react with CO₂ when the power plant and thus the carbonator enter into operation in back-up mode. For the sake of simplicity, it is considered that there is sufficient gas–solid contact time in the carbonator reactor to allow a conversion of 80% with respect to the maximum conversion attainable by the solids (X_{ave}). This high level of conversion has been demonstrated to be feasible in several large-scale calcium

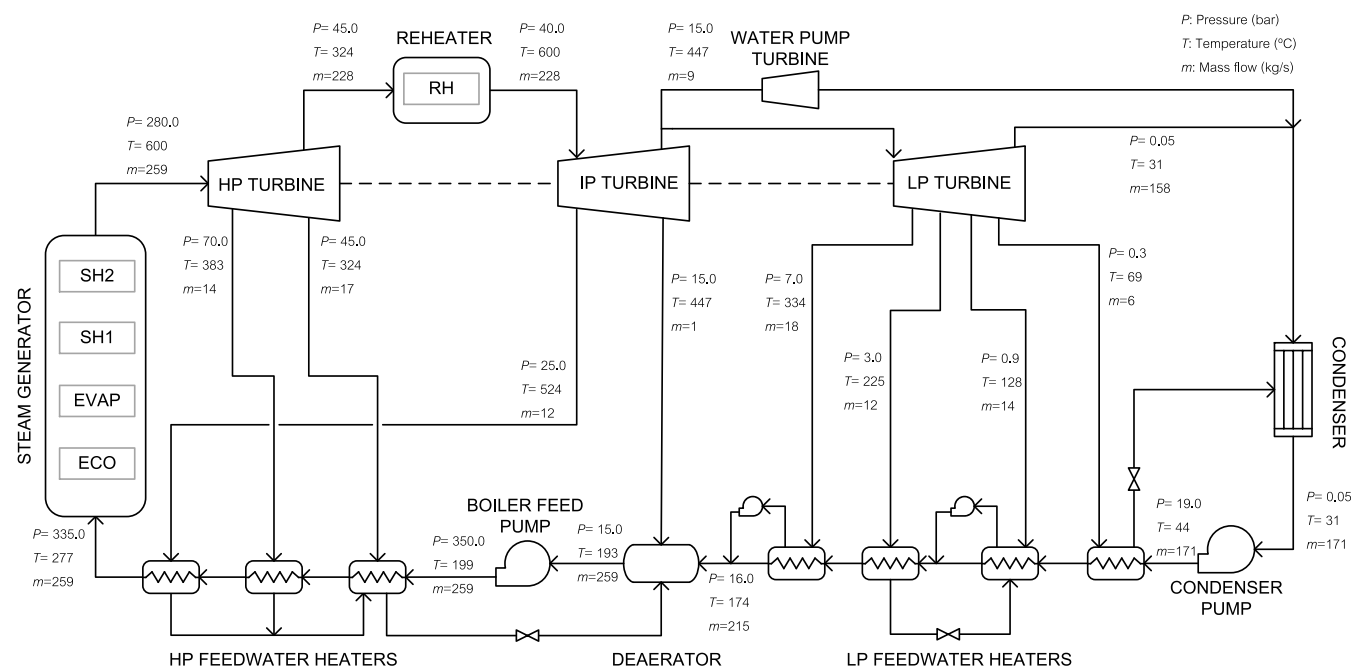


Figure 2. Reference power plant supercritical steam cycle. Pressure (P in bar) and temperature (T in °C) operation conditions. Mass flow (m in kg/s) reported correspond to a net electric power of 350 MW_e.

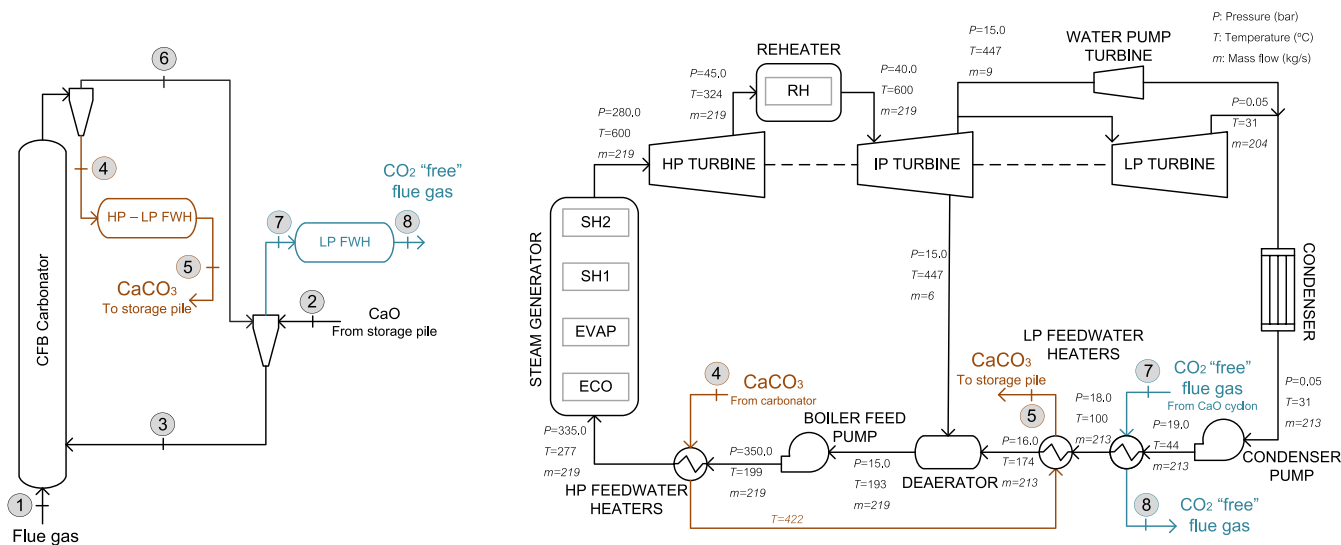


Figure 3. Carbonator block process scheme including the integration of the solids and gas streams (left) and the modified power plant supercritical steam cycle of Figure 2 where pressure (P in bar), temperature (T in $^{\circ}\text{C}$), and mass flow (m in kg/s) operation conditions for the reference case are indicated (right).

looping pilot plants where the carbonator reactor is a CFB operated with active space times of around 20–80 s.^{23,50,51}

The operation temperature of the adiabatic carbonator is directly linked to the conversion of the solids and the inlet temperature of the calcined solids entering the reactor. Several integration approaches can be put in practice to preheat the gas and solids streams entering the carbonator. These will ensure that, even for modest solid storage temperatures, standard temperatures of around 650 $^{\circ}\text{C}$ are achieved in the adiabatic carbonator during the steady-state operation mode thanks to the exothermic carbonation of CaO. In this way, the high-temperature heat available in the CO_2 “free” gas stream can be employed to heat up the flow of CaO stored at modest temperature before it enters into the carbonator reactor. Alternatively, the hot CaCO_3 leaving the carbonator reactor can be used to preheat the flue gas. These heat-exchange operations can be carried out by means of conventional gas–solid heat transfer methods or gas–solid cyclonic suspension preheaters, such as those available in cement plants.

For this specific case with a X_{ave} of 0.35, a solids inlet temperature of 410 $^{\circ}\text{C}$ is needed for the carbonator to operate at a temperature of 650 $^{\circ}\text{C}$. This can be achieved by bringing the flue gas leaving the carbonator into contact with the solids coming from the CaO pile (as depicted in Figure 3), assuming that they have been stored at a temperature of 156 $^{\circ}\text{C}$ (please note that the CaO pile is feed from solids coming from the oxy-caliner reactor after recovering part of their heat for power production, as detailed in the following section). In comparison with the previously proposed scheme,²² this preheating step allows the storage temperature in the CaO pile to be reduced by around 330 $^{\circ}\text{C}$ if the carbonator is operating at 650 $^{\circ}\text{C}$. After this preheating step, the residual heat from the CO_2 flue gas and the heat available in the stream of carbonated solids leaving the carbonator can be integrated within the existing power plant steam cycle in the feedwater preheater sections, as schematically shown on the right-hand side of Figure 3. The heat available from the CaCO_3 stream can be used both in the high- and low-pressure feedwater heaters, thereby avoiding bleeds of steam from the high-, intermediate-, and part of the low-pressure turbines. The heat

contained in the solids can be extracted in counter-current mode in moving bed heat exchangers (similar to that proposed by Nsakala et al.⁵² for oxy-fired power plants). On the other hand, the waste heat available in the CO_2 “free” stream after the CaO preheating step can be used in the low-pressure feedwater preheater section by extracting heat in counter-current water–gas heat exchangers. In this way, all of the low pressure steam bleeds can be avoided. In both heat exchangers, the temperature of the water before entering the deaerator and the economizer has been kept constant as in the reference power plant and the outlet temperatures of the solids and gases have been calculated.

The use of heat from the gas and solids streams to heat up the water in the feedwater heaters allows the existing steam cycle to operate in two different ways, by increasing the effective power of the steam turbine or by reducing the power plant thermal input. On the one hand, the coal thermal input can be maintained at its nominal load of 777 MW_{th} and the extra carbonation heat can be used to boost the electrical power output from the power plant steam turbines. On the other hand, the coal thermal input in the back-up power plant can be reduced while producing the same electrical output of 350 MW_e as in the reference back-up power plant. One of the advantages of this second approach is that the amount of CO_2 produced during coal combustion decreases and, therefore, the size of the FlexiCaL elements and streams can be reduced. For these reasons, this second approach has been adopted in this study and the process of Figure 1 has been designed to produce the same net electrical output as in the reference power plant. Although there are some minor changes to the steam flow in the different turbine sections with respect to the reference case due to the absence of bleeds (<15%), the effect of these changes on the isentropic efficiencies of the steam turbine sections can be safely ignored and the same values of pressure and temperature as those shown in Figure 2 have been assumed.⁵³

By following this second approach, the thermal power plant input can be reduced by 12% (i.e., down to 685 MW_{th}) with respect to the reference case. As is discussed in the following sections, the net efficiency of the whole FlexiCaL system will

Table 2. Specifications of the Carbonator Block Streams (See Figure 3) for a Scenario with a Thermal Power Input to the Back-Up Power Plant of 685 MW_{th} and a $X_{\text{ave}} = 0.35$

nos	description	temperature (°C)	mass flow (kg/s)	heat available (MW _{th}) ^a	composition							
					solid (% wt)				gas (% v)			
					CaO	CaCO ₃	CaSO ₄	Ash	CO ₂	N ₂	H ₂ O	O ₂
1	flue gas from power plant	140	275.0	35.2					15.3	75.7	6.3	2.7
2	CaO from storage pile	156	267.7	32.6	94.6	0.0	4.2	1.2				
3	CaO entering carbonator	410		96.7								
4	CaCO ₃ leaving carbonator	650	323.3	216.5	56.4	39.1	3.5	1.0				
5	CaCO ₃ to storage pile	207		58.7								
6	CO ₂ “free” gas leaving carbonator	650	219.3	166.1					1.8	87.8	7.3	3.1
7	CO ₂ “free” gas to steam cycle	410		98.7								
8	exhaust CO ₂ “free” gas	161		34.4								

^aReference temperature 20 °C.

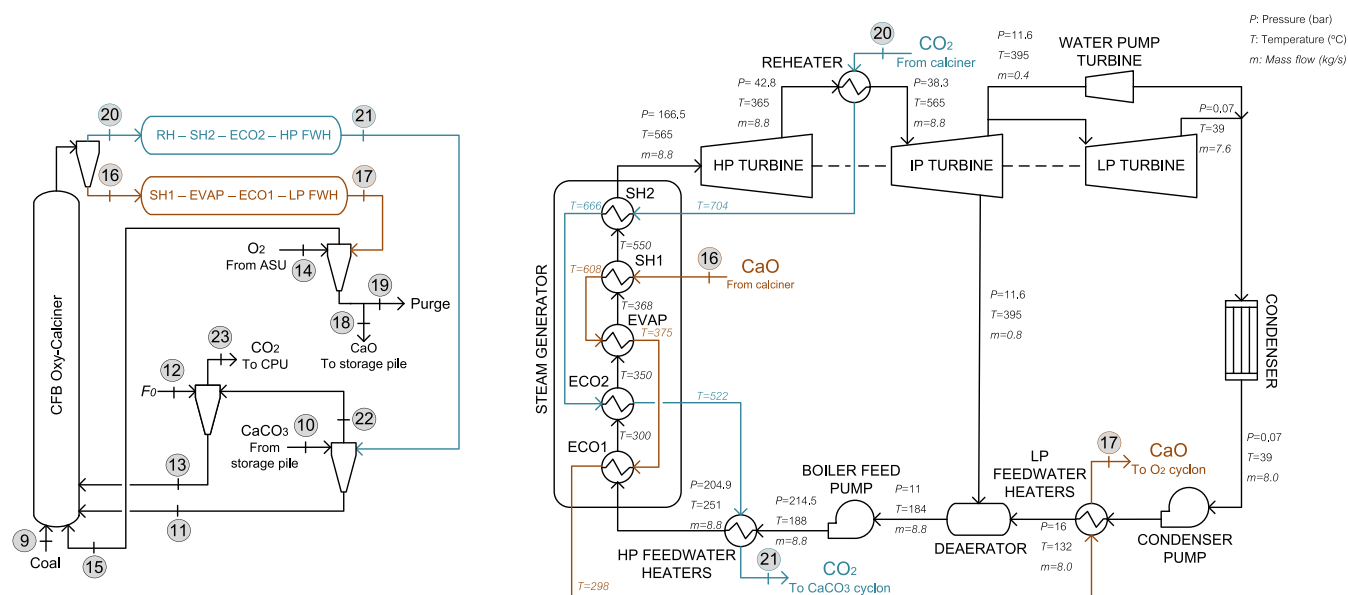


Figure 4. Oxy-calcliner block integration of the solids and gas streams (left) and the sub-critical steam cycle (right). Pressure (P in bar) and temperature (T in °C) operation conditions for the steam cycle obtained from data available in the literature.^{58,59} Mass flow (m in kg/s) reported for the reference case solved with an oxy-calcliner gross electrical output of 12.5 MW_e.

be lower due to the penalty associated with the fuel requirements in the oxy-calcliner block. This integration scheme makes it possible to transfer around 145 MW_{th} for water preheating in the steam cycle from the solids leaving the carbonator, which are cooled down to a temperature of 207 °C before being sent to the CaCO₃ pile. Meanwhile, the temperature of the CO₂ “free” gas stream is reduced to 161 °C before it is emitted, providing 61 MW_{th} for the low-pressure feedwater preheaters. As a result, a total amount of 206 MW_{th} from the carbonator block is used to produce power in the existing steam cycle. The elimination of steam bleeds from the turbine has a penalty effect on the net steam cycle efficiency (defined as the ratio between the net electrical power output and total thermal power input to the steam generator and feed water sections of the steam cycle) as this is reduced from 48% in the reference steam cycle of Figure 2 to 42% in the configuration shown in the right-hand side of Figure 3 where the steam bleeds from the turbines are avoided.⁵⁴ A summary of the main results of the balances solved for the carbonator block in the selected reference case is presented in Table 2

with the specifications of the process streams numbered in Figure 3.

3.3. Sorbent Regeneration and Thermal Integration of the Oxy-Calcliner Block.

The equipment required for the regeneration of the sorbent includes a refractory adiabatic CFB combustor-calcliner as well as an ASU that provides the required O₂ and a CO₂ compression and purification unit (CPU). The use of the solids piles allow to be operated in continuous mode. Thus, a continuous flow of carbonated solids is fed into the oxy-calcliner from the CaCO₃ pile which acts as a buffer. As a result, a more reduced flow of solids goes from the CaCO₃ to the CaO pile through the oxy-calcliner compared to the flow of calcined solids that is fed into the carbonator when the power plant is in operation. This also allows the size of the main components related with the calcination step to be reduced, as schematically shown in Figure 1; meanwhile, a modest flow of CO₂ is generated in steady state in the oxy-fired calcliner. This should facilitate both the transport and storage of CO₂ because the range of

Table 3. Specifications of the Oxy-Calcliner Block Streams (See Figure 4) for a Scenario with a Thermal Power Input to the Back-Up Power Plant of 685 MW_{th} and a $X_{ave} = 0.35$

no	description	temperature (°C)	mass flow (kg/s)	heat available (MW _{th}) ^a	composition								
					solid (% wt)				gas (% v)				
					CaO	CaCO ₃	CaSO ₄	ash	CO ₂	N ₂	H ₂ O	O ₂	
9	coal to oxy-calcliner	20	2.0	0.0	% wt (78.8 C, 4.7 H, 0.7 S, 5.8 O, 2.6 H ₂ O, 1.2 N, 6.2 ash)								
10	CaCO ₃ from storage pile	207	32.3	5.9	56.4	39.1	3.5	1.0					
11	CaCO ₃ entering oxy-calcliner	270		7.9									
12	make-up flow (F_0)	20	5.2	0.0	0.0	100.0	0.0	0.0					
13	F_0 entering oxy-calcliner	207		1.0									
14	O ₂ from ASU	20	5.6	0.0					0.0	5.0	0.0	95.0	
15	O ₂ entering oxy-calcliner	156		0.7									
16	CaO leaving oxy-calcliner	910	29.8	26.3	94.6	0.0	4.2	1.2					
17	CaO after steam cycle	183		4.4									
18	CaO to storage pile	156	26.8	3.3									
19	Purge	156	3	0.4									
20	CO ₂ leaving oxy-calcliner	910	15.3	18.2					80.9	2.5	12.1	3.5	
21	CO ₂ after steam cycle	385		6.4									
22	CO ₂ to F_0 pre-heater	270		4.1									
23	CO ₂ to CPU	207		3.0									

^aReference temperature 20 °C.

geological CO₂ storage sites broadens thanks to the reduced amount of CO₂ to be stored per year.

To satisfy the electricity requirements of the ASU, CPU, and other auxiliary units associated with the oxy-calcliner block, heat from the gas and solids streams leaving the oxy-calcliner reactor can be recovered for power generation in a separate small steam cycle. After recovery, the remaining waste heat from these streams can be further employed to preheat the gas and solids entering the calcliner in order to reduce the heat demand and fuel consumption, as below described. For the sake of simplicity, coal is assumed to be the fuel in the calcliner. However, in future scenarios with a wider availability of renewable electricity, electrolytic hydrogen and oxygen could be available (as first pointed out by Steinbeck and Dettmann⁵⁵) at a reasonable cost,⁵⁶ or other alternative biofuels could be employed to fire the small calcliner required to regenerate the sorbent.

A temperature of 910 °C is considered to completely regenerate the CaCO₃ in the oxy-calcliner reactor. Coal (78.8% C, 4.7% H, 0.7% S, 5.8% O, 2.6% H₂O, 1.2% N, and 6.2% ash, heating lower value of 32.5 MJ/kg) is burned using oxygen supplied from an ASU with a purity of 95% and in an excess of a 10%. In order to estimate the make-up flow of limestone (F_0), the equation proposed by Rodríguez et al.⁵⁷ has been used taking into account the effect of the CF on the average CO₂ flow fed into the carbonator (F_{CO_2}) as indicated in a previous work.²² This yields a $F_0/(F_{CO_2}CF)$ ratio of 0.37 which translates into an annual limestone consumption of 0.16 Mton/year.

Most of the high-temperature heat available in the streams of gas and solids leaving the oxy-calcliner reactor is used to produce power in a subcritical steam cycle, as shown in Figure 4. Pressure and temperature conditions in this cycle have been chosen in accordance with the data available in the literature.⁵⁸ To provide the required power for this subcritical steam cycle, heat is extracted from the CaO solids and CO₂-rich gas streams arriving from the oxy-calcliner at a temperature of 910 °C in

accordance with the scheme shown in the left-hand side of Figure 4. A minimum pinch temperature of 25 °C has been assumed in the different heat exchangers. In the proposed integration scheme, the heat contained in the CO₂-rich gas leaving the calcliner is used in the reheating, superheating, economizer, and high-pressure feedwater preheater sections, which allows this stream to cool down to 385 °C. The waste heat from this gas stream is then recovered to preheat the stream of CaCO₃ solids and the make-up flow of limestone respectively before they enter the oxy-calcliner. These heat exchanges steps allows the CO₂-rich flue gas stream temperature to be further reduced to 207 °C before it is sent to the CPU unit.

The heat contained in the calcined solids is also used in the superheating, evaporation, economizer, and low-pressure feedwater heater sections, as shown in Figure 4. This reduces the CaO stream temperature to 183 °C. In order to further cool down the solids and achieve the temperature needed in the CaO pile (i.e., 156 °C as discussed in Section 3.2), the waste heat can be employed to preheat the flow of O₂ fed into the oxy-calcliner, as indicated in Figure 4. More detailed specifications of the process streams numbered in Figure 4 for the oxy-calcliner block can be found in Table 3.

In accordance with previous assumptions and the integration scheme shown in Figures 3 and 4, a thermal input to the oxy-calcliner as low as 66 MW_{th} is calculated. This represents a fraction of only 8% with respect to the total thermal capacity (including that of the back-up power plant). This compares favorably with conventional CaL systems that require a typical thermal capacity of around 45–50% for the oxy-fired calcliner.³¹ Another benefit of the system proposed is that a continuous flow of 0.31 kmol/s of CO₂ is produced in the calcliner in contrast with the large flow of CO₂ captured from the power plant emitted during operation periods at a rate of 1.41 kmol/s. Consequently, the size of the CPU needed is reduced by almost 78%.

With the integration of Figure 4, a gross efficiency of 40% is calculated for the oxy-calcliner block, resulting in a gross

electrical power output ($P_{e, \text{gross, oxy-calciner}}$) of 12.5 MW_e. On the basis of typical assumptions for power consumption in the main elements of the oxy-calciner block (i.e., 5% of the gross power output in the auxiliaries⁶⁰ and 200 kWh_e/t_{O₂} and 120 kWh_e/t_{CO₂} in the ASU and CPU units, respectively⁶¹), an internal consumption of around 11 MW_e is estimated for the oxy-calciner block. Thus, it can be assumed that no net power is produced by the steam cycle associated with the calciner. The power delivered to the grid comes only from the steam cycle associated with the back-up power plant and carbonator blocks.

3.4. Global Process Performance. Accordingly to the discussion of the previous sections, the net efficiency of the FlexiCaL system (η_{FlexiCaL}) as depicted in Figure 1, exporting power only during a fraction of the time CF and considering that the net power output is zero from the oxy-calciner thermal input, can be calculated using the following equation

$$\eta_{\text{FlexiCaL}} = ((P_{e, \text{net}} \text{CF}) / (P_{\text{th, power plant}} \text{CF} + P_{\text{th, oxy-calciner}} - P_{\text{th, F}_0})) \cdot 100 \quad (1)$$

where $P_{e, \text{net}}$ is the net electrical power produced in the system and $P_{\text{th, power plant}}$ and $P_{\text{th, oxy-calciner}}$ are the thermal inputs to the power plant and the oxy-fired calciner, respectively, and reported in the previous sections. The thermal input associated with the calcination of the fresh limestone ($P_{\text{th, F}_0}$) can be discounted considering the benefits from the potential use of the CaO-rich purge (around 0.1 Mton/year) in a cement plant or another large-scale CaO consumer. From eq 1, a η_{FlexiCaL} of 28% is calculated; meanwhile, the net power efficiency of the reference power plant without CO₂ capture was of 45%. Despite this modest value it should be emphasized that the energy penalty related to the regeneration of the sorbent is applied during the whole operating period, while the back-up power plant is able to deliver a net power output of 350 MW_e to the grid when it is required to enter into operation and capturing the CO₂ in the flue gas with a 90% efficiency.

Regarding the capacity of the CaO and CaCO₃ piles depicted in Figure 1, around 20 800 and 25 100 m³, respectively, (assuming a bulk density of 1000 kg/m³) are needed per operation day of the back-up power plant. As mentioned above, for the integration scheme proposed, the CaO and CaCO₃ are stored at temperatures of around 150–200 °C. This facilitates the solid handling and storage operations by making it possible to use equipment normally employed in other processes (e.g., in the cement industry).²²

Finally, the main process parameters of the reference case solved in the previous sections have been summarized in Table 4. The reference case here studied shows that the FlexiCaL process configuration represents a promising CO₂ capture system for retrofitting and extending the lifetime of amortized back-up fossil fuel power plants forced to operate under extremely low-capacity factors.

4. CONCLUSIONS

Calcium looping system relying on large reservoirs of low-temperature CaCO₃ and CaO can be used to capture the CO₂ emitted from existing back-up coal power plant forced to operate under extremely low CFs. One of the main features of the FlexiCaL process is the use of existing equipment in the power plant to minimize the need of additional equipment associated with the CO₂ capture and to avoid waste of capital.

Table 4. Summary of the Global Performance of the FlexiCaL Reference Case Studied in This Work

description	unit	value
net electrical power output, $P_{e, \text{net}}$	MW _e	350
power plant thermal input, $P_{\text{th, power plant}}$	MW _{th}	685
power recovered from carbonator block to the power plant steam cycle, $P_{\text{th, carbonator}}$	MW _{th}	206
oxy-calciner thermal input, $P_{\text{th, oxy-calciner}}$	MW _{th}	66
oxy-calciner block gross electrical power output, $P_{e, \text{gross, oxy-calciner}}$	MW _e	12.5
net efficiency of the FlexiCaL system (eq 1), η_{FlexiCaL}	%	28
storage silos capacity, V_{storage}	m ³ /day	20 800–25 100

The thermal integration exercise carried out reveals that it is possible to maintain the net power output of the back-up plant with capture, by sacrificing the efficiency of the system when regenerating the CaCO₃ to CaO. This is because the proposed system includes an adiabatic carbonator reactor that can follow the power plant operation periods, whereas the small-scale oxy-calciner block produces power for the autoconsumption in the air separation and compression units (i.e., zero net power exports). A reference case solved for a CF of 0.1 indicates that 12% of the energy during back-up power operations comes from the heat recovered from the carbonator block when using the feedwater preheater sections of the existing power plant steam cycle. For a reference output of 350 MW_e, and assuming a reasonable solids storage activity (X_{ave} of 0.35), a thermal input of just 66 MW_{th} (which represents only 8% of the total thermal capacity of the system) is required in the oxy-calciner block to achieve the complete regeneration of the carbonated solids, storing the solids at moderate temperatures of around 150–200 °C. A large volume of solid reservoirs is needed (around 20 000–25 000 m³ of bulk solids per day operating at full load), but the very low specific cost of CaO and CaCO₃ materials could make this process attractive in future scenarios where capture of CO₂ from back-up power plants may be needed.

■ ASSOCIATED CONTENT

Supporting Information

The Supporting Information is available free of charge at <https://pubs.acs.org/doi/10.1021/acsomega.9b03552>.

Composite curves of the carbonator and oxy-calciner gas and solids streams integration with their corresponding steam cycles (PDF)

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Notes

The authors declare no competing financial interest.

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NOMENCLATURE

CF	capacity factor
E_{calc}	calcination efficiency
E_{carb}	CO ₂ capture efficiency
F_0	make-up flow, mol/s
F_{CO_2}	average CO ₂ flow fed into the carbonator, mol/s
m	mass flow, kg/s
P	pressure of steam cycle streams, bar
$P_{\text{e,net}}$	net electrical power output, MW _e
$P_{\text{e gross,oxy-calciner}}$	oxy-calciner block gross electrical power output, MW _e
$P_{\text{th,carbonator}}$	power recovered from carbonator block to the power plant, MW _{th}
P_{th,F_0}	thermal input associated with the calcination of F_0 , MW _{th}
$P_{\text{th,oxy-calciner}}$	oxy-calciner thermal input, MW _{th}
$P_{\text{th,power plant ref}}$	power plant thermal capacity, MW _{th}
$P_{\text{th,power plant}}$	power plant thermal input, MW _{th}
T	temperature, °C
T_{calc}	oxy-calcination temperature, °C
T_{carb}	carbonator temperature, °C
V_{storage}	storage silos capacity, m ³ /day
X_{ave}	average CO ₂ carrying capacity
η_{FlexiCaL}	net efficiency of the FlexiCaL system according to eq 1, %
$\eta_{\text{power plant}}$	net efficiency of the reference power plant without CO ₂ capture, %

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