16th International Conference on Greenhouse Gas Control Technologies, GHGT-16

23rd -27th October 2022, Lyon, France

Preliminary analysis of Calcium Hydroxide based Calcium Looping for low capacity factor coal-fired power plants

Fierro José J.[a], Romano Matteo C.[a], Astolfi Marco[a]*

[a]*Politecnico di Milano, Department of Energy, via Lambruschini 4, 20156 Milan, Italy*

Abstract

Calcium looping (CaL) is a promising technology for power plants having a low capacity factor thanks to the possibility to store solids and to decouple the operation of carbonator and calciner islands. The power plant considered in this work is a coal fired Ultra Super Critical steam Rankine cycle, to be used for back-up power generation in electric grids dominated by intermittent-renewables, having a capacity factor equal to 10% that would discourage the adoption of a capital intensive CO_2 capture plant sized on the full power plant capacity. A CaL process using $Ca(OH)_2$ as a sorbent gives the possibility to simpler and lower cost entrained flow carbonation reactor design, thanks to the fast kinetic that requires a short residence time (<5s) to achieve high CO_2 capture efficiency.

In this work, the preliminary results of the energy management of a $Ca(OH)_2$ calcium looping system are presented. Aspen Plus V11 is used for the overall system simulation while a 1-D model is adopted to provide a sizing of the carbonator. Results confirm that $Ca(OH)_2$ sorbent allows the design of a 60 m long adiabatic carbonator. The SPECCA index is equal to 7.8 MJLHV/kgCO_2 with a CO_2 capture efficiency of 95% and accounting for the negative emissions for the captured CO_2 from biomass oxy-combustion in the calciner. Despite the SPECCA indicates a high energy efficiency penalty compared to conventional CO_2 capture technologies, the study shows the conceptual technical feasibility of the proposed concept, to be validated experimentally and with economic analysis.

Keywords: CCS, CO_2 Capture, Calcium looping, $Ca(OH)_2$, Backup power plant

---

* Corresponding author. *E-mail address:* marco.astolfi@polimi.it

Nomenclature

CCS	Carbon Capture and Storage
CaL	Calcium Looping
LCOE	Levelized cost of electricity
PBT	Payback time
SPECCA	Specific Primary Energy Consumption for CO_2 Avoided
WHR	Waste Heat Recovery

1. Introduction

Agreements on global environmental policies like the one in Paris recently revisited and reinforced during the COP26 in Glasgow, keep in sight the actions to limit the increase in global temperature to 1.5°C above pre-industrial levels by 2100. To achieve such goals, countries need to take concerted and immediate action to deliver on their commitments, which consist among others on phasing down coal power generation, halting and reversing deforestation, speeding up the switch to electric vehicles, and reducing methane emissions in the pursuit of carbon neutrality by 2050 [1]. Power generation, nowadays heavily dependent on unabated use of fossil fuels, is not sustainable on long-term, and migration to cleaner and renewable energy sources is mandatory. Such a substantial paradigm shift poses new challenges in grid stability, due to the inherently fluctuating nature of renewables, until large-scale and low-cost energy storage systems are fully available [2],[3]. Consequently, power generation is expected to go through a gradual transition where fossil fuels will still play an essential but decreasing role moving from baseload to mainly load-following and finally to backup units in a highly decarbonized future [4]. Coal is widely used in large-scale, energy-intensive applications because of its high availability and low price, accounting for 27.2% of total global primary energy consumption in 2021 [5]. Still, it is also the fuel with the highest environmental impact, making carbon capture and storage (CCS) a fundamental tool to reduce emissions from power generation for the next decades. Among the different solutions for CCS, Calcium looping (CaL) is a promising technology for the retrofitting of existing fossil fuel-based power plants and industrial applications. In a conventional CaL scheme, in the first reactor, the sorbent's carbonation, traditionally CaO, is carried out with CO_2 from flue gas producing $CaCO_3$. Then, the $CaCO_3$ is calcined with oxygen, generating back CaO and a stream rich in CO_2 for storage or direct utilization. For base-load applications, the two reactors are sized on the nominal flue gas CO_2 content, and they work with the same capacity factor of the fossil fuel power plant or the industrial process. On the contrary, in discontinuous or variable load applications, CaL gives an additional benefit allowing for the easy decoupling of the carbonator and calciner islands thanks the use of an inexpensive and easily storable solid sorbent that would allow for a downsizing of calciner reactor and auxiliary equipment. This feature is investigated by introducing an intermediate energy/solids storage for partial load operation [6],[7], leading to a reduction in the cost of the CO_2 avoided between 16-26%, and LCOE reduction by 4-5%, compared to the reference CaL plant without solids storage, for representative medium and low-capacity factor scenarios respectively. Similar works have been proposed in [8], where the decoupling of the carbonator and calciner island with intermediate energy storage allows for a reduction in the cost of electricity when compared to standard CaL forced to operate at low capacity. Thermal integration is presented in [4], where the additional heat recovered from the carbonator products is used in the power plant steam cycle, reducing the size of the carbonator block by 12%, while the heat obtained in the calciner island is used to power a smaller steam cycle for auto consumption. In [9], it is exploited the advantage of intermediate energy storage between carbonator and calciner to reduce the energy penalty on CaL or for novel applications like concentrated solar power [10],[11]. In these studies, carbonation and calcination are carried out in a circulating fluidized bed reactor to meet the required reaction times >30~s [12]. Additionally, the CO_2 capture performance of the carbonator is affected by the cumulative deactivation of the sorbent, and loss of porosity, due to sintering of the CaO grains over carbonation/calcination cycles [13]. Efforts have been made to improve the performance of CaL in nominal and part-load operations. i.e. optimizing the calcination environment, reducing calcination temperature, modifying the physical properties of the sorbent or improving the reactivation mechanisms [14],[15]. The above-mentioned aspects can strongly limit the application of conventional CaL schemes and components for fossil fuel-based power plants operated as backup units and characterized by very low capacity factors where the need to limit capital investment cost is of paramount importance. The substitution of the sorbent from CaO to $Ca(OH)_2$ is considered a promising option in this context thanks to the faster kinetics for the carbonation reaction, which would allow decreasing the required residence time below 5s, to achieve high CO_2 conversion, thus, allowing the utilization of simpler and cheaper alternatives to typical fluidized bed reactors like once through entrained flow, and cyclonic reactors [12]. However, the use of $Ca(OH)_2$ as sorbent also implies challenges and technical uncertainties mainly related to (i) the solids handling and their heat recovery integration in the power plant, (ii) the adoption of nonconventional carbonator reactors, and (iii) the need of an intermediate hydrator reactor to produce the $Ca(OH)_2$ from the CaO released from the calciner.

This paper, which is part of the BackCap RFCS European project [16], focuses on the analysis of the implementation of a novel $Ca(OH)_2$-based CaL technology for the retrofitting of low-capacity factor coal-fired power plants in the frame of the BackCap project. The performance of each subsystem is assessed through numerical simulations with commercial software and proprietary routines. The CaL strategy is modelled in Aspen Plus V11 with simple equilibrium calculation for the carbonator (Section 2). Preliminary results with a suitable energy management configuration description are presented and discussed (Section 3). Finally, a discretized 1D model based on [17] has been implemented for the carbonator to solve momentum and energy balances (Section 4).

2. Methodology

Figure 1 depicts the calcium looping system proposed in this study for CO_2 capture from an existing coal power plant. It encompasses two main sections: the carbonator island, where CO_2 is removed from power plant flue gas using $Ca(OH)_2$ as sorbent, and the calciner island, where high purity CO_2 is released and compressed for transport to the storage site. The two islands are connected by a loop of solid sorbent that can be stored, allowing for an effective decoupling of the operation of the carbonator and the calciner reactors and auxiliary equipment, eventually reducing the capital cost of the calciner island.

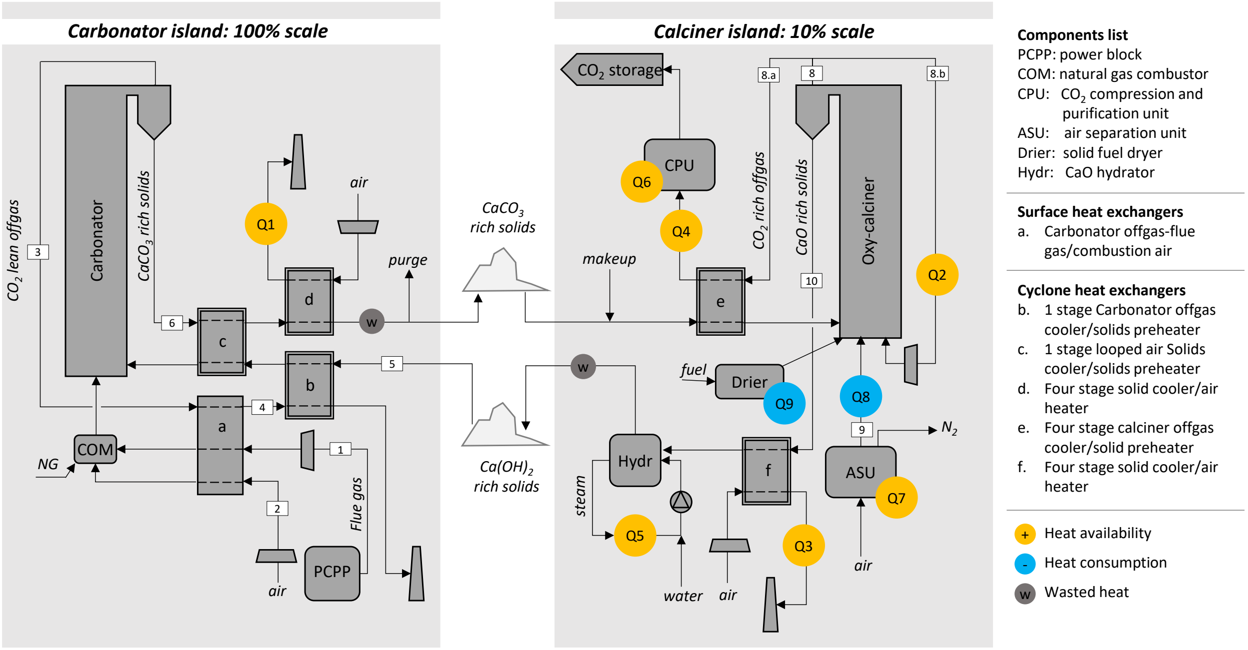

Figure 1. Block-flow diagram of the CaL strategy for CCS in back-up power plants

In the carbonator island, the main goal is to maximise the exploitation of the heat available from the products to preheat the reactants and limit the consumption of additional fuel. The flue gas (1) released after SO_2 removal from the existing pulverized-coal power plant (PCPP) and the combustion air (2) are preheated cooling a fraction the lean CO_2 carbonator offgas (3). This process can implement a surface heat exchanger (components a) like three-way Ljungström or tubular-type air preheaters. Cooled CO_2-lean offgas (4) can be further cooled down by preheating $Ca(OH)_2$ available at ambient temperature from storage (5), adopting a single-stage cyclone preheater (b). CO_2 lean offgas is then released at the stack. $Ca(OH)_2$ preheating is completed via indirect heat transfer from hot carbonated solids (6) to colder sorbent, with air as heat transfer fluid circulating between two double stage cyclone contactors (c). The inlet temperature of $Ca(OH)_2$ entering the carbonator is 325°C which is the maximum preheating temperature to avoid premature $Ca(OH)_2$ dehydration and preserve high carbonation kinetics [18] The proposed carbonator heat management maximises flue gas, air, and solid preheating, reducing natural gas (NG) consumption required to ensure the carbonator energy balance and to achieve a mixing temperature at the carbonator inlet of 550°C, sufficient to initiate the carbonation reaction with fast kinetics. $CaCO_3$-rich solids from the carbonator are eventually cooled in a four stage cyclone cooler (d) down to 35°C, heating ambient air. Residual sensible heat is wasted by natural convection during long-term storage. In contrast, heat available from hot air (Q1) can be recovered, for example in the existing power plant feedwater preheating, leading to a higher power production due to reduced steam bleeding from the turbine.

In the calciner island, the main goal is to reduce the amount of fuel (woody biomass in this case) burned in the calciner in oxy-fuel condition, through an efficient preheating of the solid sorbent. CO_2 rich off gas (8) are split into two streams: the first one (8.a) is used to preheat the sorbent in a four stage cyclone preheater (e) while a smaller amount (8.b) is initially cooled down to 500°C (Q2) and sent back to the calciner to keep oxygen content at calciner inlet equal to 50%. Oxygen (9) is produced by an Air Separation Unit (ASU) from ambient air and then heated up to 150°C (Q8) before entering the calciner. Biomass is dried (Drier) (Q9) with hot air in a belt drier. Hot CaO solids (10) at 920°C are cooled down to 35°C with a four stage cyclone cooler (f), where the air is heated up. Cold CaO eventually enters the hydrator (Hydr) reactor, where water is provided for the conversion of CaO to $Ca(OH)_2$. The hydration process is carried out at ambient pressure, and the heat of the reaction is used to evaporate the water excess. Steam released from the hydrator is condensed and recirculated while hot $Ca(OH)_2$ at around 100°C is stored. Makeup and purge mass flow rates are controlled to have an ash content below to 5% at the hydrator inlet. The residual sensible heat of $Ca(OH)_2$ is wasted while a large amount of heat is available from different sections of the calciner island (Q2-Q7). At the same time, only a small fraction is required for drying (Q9) and oxygen preheating (Q8) processes. This allows considering the use of heat recovery systems for power production to improve the plant efficiency, as explained in detail in section 3.

The power generation plant used as a reference in this work consists of an air combustion boiler burning low sulfur bituminous coal (LHV = 25.17MJ/kg and 14.15% ash content), coupled with a pulverized coal power plant (PCPP) with an Ultra Super Critical (USC) steam cycle as described in [6]. This PCPP has been assessed for CCS applications at design and part load regimes in [6],[7] and it is characterized by a total thermal input equal to 1676.57 MW_{LHV}, a net electric power output of 747.2 MWe and a total CO_2 emissions of 162,9 kg/s. In this work, it is considered operating at a backup regime, with a capacity factor of 10% on an 8030h/y availability time.

Regarding the numerical model developed in Aspen Plus V11, a stoichiometric reactor was chosen for the carbonator. Following a two-step reaction, as shown in equations 1-2, full dehydration of $Ca(OH)_2$ is followed by the carbonation of the nascent CaO up to a 70% conversion, as suggested by [12], resulting in a slightly exothermic process (around -21.6 kJ/mol at considered carbonator operative temperature). The calciner was calculated assuming complete combustion and $CaCO_3$ decomposition. These two reactors are considered adiabatic and are linked with a heat stream in such a way that the heat produced in the combustion reaction is introduced in the calcination one [19]

$$Ca(OH)_2 \rightarrow CaO + H_2O \qquad [+99.9; 97.3]\, kJ/mol \quad for\, [800; 900]\ K \qquad (1)$$

$$CaO + CO_2 \rightarrow CaCO_3 \qquad [-172.5; -170.9]\, kJ/mol \quad for\, [800; 900]\ K \qquad (2)$$

The system also includes an air separation unit (ASU) for oxygen production and a CO_2 compression and purification unit (CPU).

$Ca(OH)_2$ is produced trough hydration of CaO with liquid water. Water is feed in excess to control the temperature of the reaction at around 100°C. The excess water is vaporized and condensed back and returned into the process, so that makeup water is only needed to compensate the reacting portion.

The most significant assumptions used to estimate the energy balance of the CO_2 capture process are listed in Table 1

Table 1 - Assumptions for the CaL system components.

Carbonator island		**Calciner island**	
Inlet temperature [°C]	550	Outlet temperature [°C]	920
CO_2 capture efficiency [%]	95	Oxygen concentration in oxidant stream [%vol]	50%
FCa/FCO_2 (Feed)	1.36	Oxygen concentration in CO_2-rich gas [%vol]	5%
$F0/FCO_2$ (Makeup)	0.33	Recycle gas temperature [°C]	500
Auxiliaries			
Cyclones efficiency on solids [%]	95%	Oxygen purity [%vol]	95%
Fans isentropic efficiency [%]	80%	ASU Electric consumption [kWh/tO_2]	160
Fans electric-mechanical efficiency [%]	94%	CPU Electric consumption [$kWhe/tCO_2$]	115.8

3. Results

This section presents the results of simulation in nominal steady-state conditions, considering the overall energy balance, the heat management on the carbonator and calciner island, and the power output theoretically available from the heat released from the process. Results refer to a capacity factor of the coal-fired power plant equal to 0.1. The size of the calciner island is 10% of the size of the carbonator island. Figure 2.a depicts the heat management of the carbonator island. Here, two main aspects need to be highlighted. First, the carbonator offgas (red line) is used to preheat both combustion air for the natural gas burner and the flue gases coming from the power plant (blue lines) up to maximum temperature (600°C), resulting from the assumed minimum pinch point temperature difference of 50°C (the two processes are graphically split although they are likely carried out in the same heat exchanger). The design of this component is challenging because of the high temperatures involved, the low heat transfer coefficient (ambient pressure gases) and the very high volumetric flow rates and duty. Non-conventional high-temperature three-way Ljungström heat exchangers or tubular type preheaters can be considered, to be carefully designed considering the unconventionally high temperatures. After $Ca(OH)_2$ preheating, the carbonated solids are cooled down in a four-stage cyclone preheater that indirectly heats air up to 209°C, recovering 177 MW. Cooling of solids by direct contact with air is considered preferable compared to a tubular heat exchanger as the small particle size of the solids make them hardly fluidizable. The recovered heat can be transferred to the existing power plant for feedwater preheating, partially or completely closing the steam bleedings from the LP turbine and increasing the total power output. This option is considered preferable compared to installing a new dedicated waste heat recovery (WHR) power plant that will operate with very low capacity factor, involving long pay back time. On the other hand, this design involves difficulties related to the increase of the volumetric flow rate in the low-pressure turbine, that will thus operate with lower efficiency. This work estimates the additional power by adopting a second law efficiency equal to 55% of the exergy variation available from hot air cooling. Thus, the resulting estimated power output from the carbonator island equals 21.2 MW. Additional consideration regards the very small impact of the last two stages of the four stage cyclone preheaters (d and f) in term of heat recovery: the impact of a reduced number of stages from technoeconomic perspective will be investigated in next step of this study.

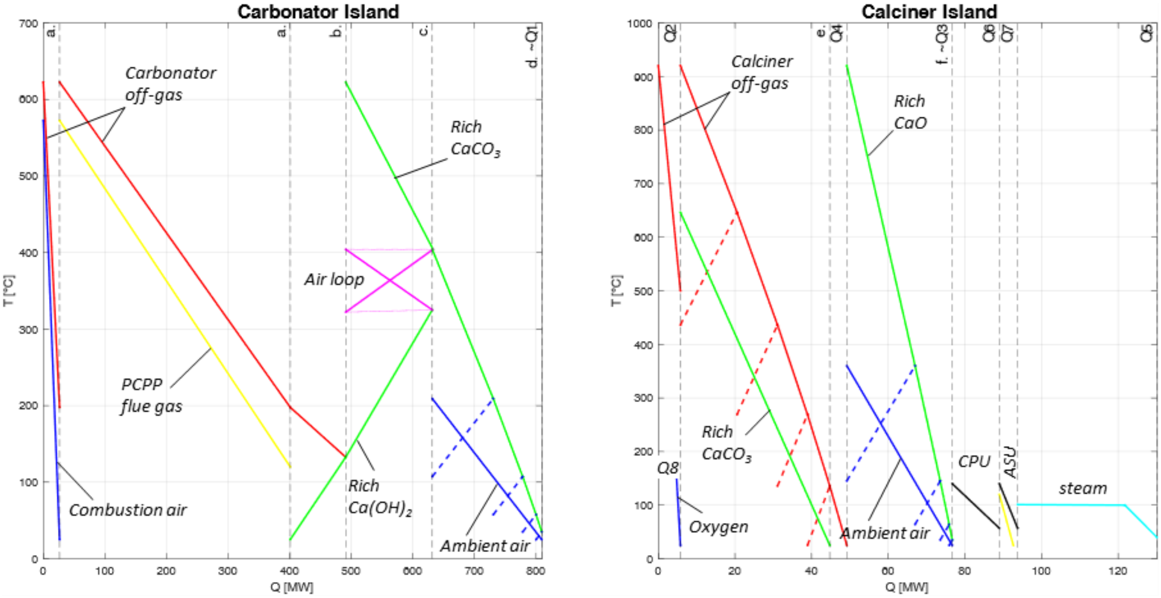

Figure 2. Temperature-heat diagram for the Carbonator Island.

Figure 2.b depicts the temperature-heat diagram of the calciner island. Heat is recovered from the calciner off gas for sorbent preheating, while hot air for the biomass drier is heated by exploiting waste heat from ASU compressors intercoolers. Oxygen is preheated with part of the thermal power available from the recirculated offgas. A large amount of heat is still available from different sections, namely: (i) high-temperature heat (above 150°C) is available from recirculated off gas (5.8 MW between 920-500°C) and from hot air heated up by solid cooling in the four-stage

cyclone preheater (g) (27.5MW between 352-25°C); (ii) low temperature heat (below 150°C) is available from final off-gas cooling, from CPU and ASU intercoolers (12.3MW and 4.8MW between 140 and 57°C), from hot air (15.25 MW between 150°C and 25°C) and from recirculated steam in the hydrator section (28 MW mostly at 100°C). For the calciner island, the installation of WHR units is highly suggested considering the base-load operation. High-temperature heat can be exploited with an advanced organic Rankine Cycle (ORC) or a supercritical CO_2 cycle (sCO_2) power plant [20], that can reach a conservative efficiency value of around 20%, while low-temperature heat can be exploited only with geothermal-like ORC with an efficiency of around 5% [21]. A final option is to design a two-evaporation level ORC as already proposed by some manufacturers [22]. A steam cycle is not recommended in this case because the overall thermal input is too small to design a cycle with sufficiently high efficiency. The estimated power output from the calciner island is 8 MW. Table 2 reports the power consumption associated with all the auxiliary equipment in the system. Power consumption of fans is calculated considering a pressure drop equal to 0.5 kPa in the surface heat exchangers, 1 kPa in the riser and 1 kPa in the cyclones. Solid conveyors and water pump consumption is neglected. The energy balance is calculated considering the capacity factor of 10% for the coal-fired power plant, resulting in an annual net energy output of 525.4 GWh (-12.4% with respect to the reference PCPP, due to the electricity consumed in the calciner island) and consumption of additional fuel (368.4 TJ of natural gas and 3384.8 TJ of biomass per year). Overall system results are reported in Table 3: the final CO_2 emissions are equal to -258.7 kton (vs. 471 kton of the PCPP and the 491.4 kton of reference system considering also NG combustion), leading to specific emissions of -492.4 $kgCO_2$/MWhe annually, while overall system efficiency drops from 44.56% to 21.99%. The Specific Primary Energy Consumption for CO_2 Avoided, SPECCA is calculated as in equation 4, where HR is the heat rate of the plants, expressed in [MJ_{LHV}/MWhe] and E is the CO_2 emission rate, expressed in [kg_{CO2}/MWhe]. HR values for the reference system is calculated considering electrical efficiency of coal, natural gas and biomass power plant equal to 44.5%, 60% and 35% respectively.

$$SPECCA = \frac{(HR - HR_{REF})}{(E_{REF} - E)} \qquad (4)[23]$$

The obtained SPECCA for the BackCap plant with the given assumptions is equal to 7.77 $MJ_{LHV}/kgCO_2$. This value is considerably higher than ~2.1-2.3 $MJ_{LHV}/kgCO_2$ obtained for conventional calcium looping processes without sorbent storage or with high temperature sorbent storage [7]. Such large difference is due to the need of reactants preheating caused by the long term ambient temperature storage, the need of additional natural gas combustion caused by the lower heat of reaction of $Ca(OH)_2$ than CaO and the resulting lower efficiency of the WHR units.

Table 2. Power balance of the PCPP + CaL plant.

	PCPP	**Carbonator**	**Calciner**
Capacity Factor [%]	100%	100%	10%
Operating time [h/y]	803	803	8030
Fuel consumption [kg/s]			
Coal [kg/s]	66.61	-	-
Biomass [kg/s]	0	-	8.66
NG [kg/s]	0	2.56	-
CO_2 from fuel [kg/s]	162.9	7.09	9.80
Power balance [MW]			
Fuel LHV [MW_LHV]	1677	127.4	117.1
Heat available [MW]	-	177.86	24.2 >150°C 62.3 <150°C
Emitted CO_2, kg/s	162.9	8.50	-9.80
Electric Power Balance [MW]			
Fans consumption [MW]		-21.80	-0.15
Power available [MW]	-	21.20	4.84 >150°C 3.11 <150°C
ASU [MW]	-	-	-4.8
CPU [MW]	-	-	-12.3
Net power output [MW]	747.2	-0.61	-9.23

Table 3. BackCap key performance indicators.

	Ref	**BackCap**
Total energy from fuel [TJ/year]	8599.8	
Coal	4846.6	
NG	368.4	
Biomass	3384.8	
Total energy output [GWh/year]	990.5	525.4
Coal (efficiency: 44.56%)	600.0	-
NG (efficiency: 60%)	61.4	-
Biomass (efficiency: 35%)	329.1	-
Total CO_2 emission [kton/year]	491.4	-258.7
Coal	470.9	24.6
NG	20.5	
Biomass	-	-283.28
Average Efficiency	41.46%	21.99%
Specific CO_2 emission [$kgCO_2$/MWhe]	496.1	-492.4
Heat rate [MJ/MWhe]	8682.5	16368.1
SPECCA [MJ_LHV/$kgCO_2$]		7.77

4. Entrained flow Carbonator design

The use of a conventional circulating fluidized bed reactor is inadequate for this application because of the small sorbent particle size (~5 μm). Replacing the relatively large particle (>100 μm) CaO sorbent with small particle $Ca(OH)_2$, which exhibits faster reaction kinetics [12], has profound implications for reducing the needed time of reaction, allowing to design of the carbonator as an entrained flow reactor, that is simpler than a CFB reactor for at least two reasons. First, a carbonator designed for $Ca(OH)_2$ does not require a heat transfer surfaces in the reactor because of the contextual presence of dehydration and carbonation reactions that leads to a slightly exothermic process. Second, the fast kinetic allows avoiding the recirculation of solids, simplifying the system operation and avoiding using a loop seal for sorbent split. Finally, the use of a CFB reactor would be excluded in any case, considering that the very small diameter of $Ca(OH)_2$ precludes fluidization. Such aspects lead to the possibility of limiting the carbonator cost [13],[14], which is decisive for low capacity factor applications.

For a preliminary design of the carbonator, a 1-D model of an entrained flow carbonator has been adapted from [17] solving the coupled momentum, mass and energy balances for the solids and gases while considering the kinetics of the reaction with $Ca(OH)_2$ [12]. The carbonation reaction of $Ca(OH)_2$ encompasses two steps consisting in $Ca(OH)_2$ dehydration to CaO, modelled with a simplified shrinking core model, followed by the instantaneous CaO carbonation reaction [12]. The temperature of the reactants (T) is estimated as the average temperature of solids and gases in each discretized volume. For the sake of simplicity, the small amount of CaO entering with the sorbent (due to cyclone preheater stage c. in Figure 1 or non-completed hydration) is considered inert since it may react with CO_2 but with a much slower kinetic (random pore model [24]) and this would not affect the overall results. The adiabatic reactor is discretized only in a longitudinal direction, considering the gradient of composition and temperature negligible in the radial direction.

Figure 3 depicts the main results of the simulation reporting the trend of different quantities vs time. Figure 3.a reports the temperature diagram, showing that the solid and the gas reach the mixing temperature on timescales of milliseconds, thanks to the very high heat transfer area due to the small particle diameter and assumed Nu=0.1 [25]. Then the temperature of solid and gas slightly rises due to the slightly exothermic overall reaction. Figure 3.b depicts the conversion of $Ca(OH)_2$ to CaO (X_{CaO}), which is limited to 1 (full dehydration) and the conversion of $Ca(OH)_2$ to $CaCO_3$ (X_{CaCO3}) limited to conversion of 0.7 because of deactivation due to sintering and layering of $CaCO_3$ that inhibits further reaction in the grain [13],[18]. Full dehydration is obtained after around 3-4 seconds, which is also the reaction time required for reaching a carbon capture efficiency of 95% (Figure 3.c) which is limited by the amount of sorbent fed to the carbonator. Figure 3.d reports the molar flow rate of the different chemical species.

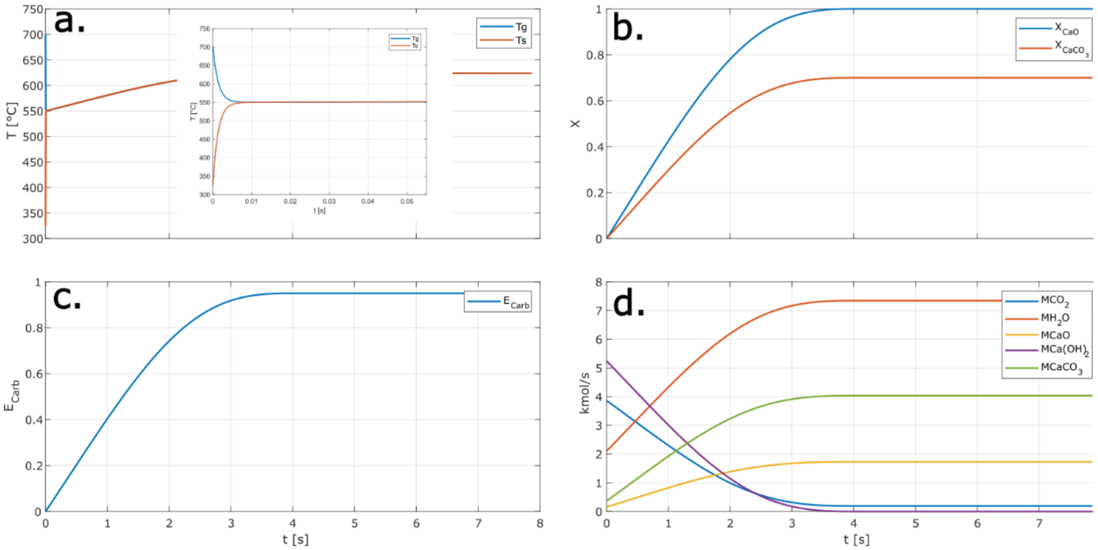

Figure 3. All the variables plotted in this chart are reffered to time of reaction [s]. a. Temperature of solids and gases [°C]. b. Conversion of $Ca(OH)_2$ and CaO. c. CO_2 capture efficiency. d. Molar flows of the reacting species [kmol/s].

Adopting an initial gas velocity of 12 m/s and considering a contact time of 4 seconds, results in a carbonator length of around 50 meters and an equivalent cross-section diameter of around 15m. It has to be noted that it may be preferable to adopt different reactors in parallel with a smaller diameter, allowing for better control of the process fluid dynamics during power plant part load operations when flue gas mass flow rate decreases. Provided that reactors are adiabatic, the adoption of parallel reactor tubes does not affect the calculation of the 1D model.

5. Conclusions

In this paper, the preliminary results of a CaL strategy for coal-based power plants operating as backup in grids dominated by intermittent renewables were presented. The carbonator island and the calciner island are operated separately and with different capacity factors (CF). Thus, the carbonator operates following the power plant operation (CF=10%), while the calciner continuously (CF=100%). The design of a compact and simple entrained flow carbonator reactor for the carbonator has been verified and confirms the potential of using $Ca(OH)_2$ as sorbent. The heat available in the carbonator and in the calciner is used to generate additional power, 21.2 MW and 8 MW, respectively that are however not sufficient to balance the auxiliaries consumptions mainly due to ASU and CPU compressors. Overall energy output decreases (-12.4%) with a strong loss of efficiency (-50%). The calculated SPECCA index is 7.8 MJ_{LHV}/kg_{CO2}, considering negative emissions from biomass oxy-combustion in the calciner. This value is higher than in plants with conventional CaL systems, where SPECCA=2.1-2.3 MJ_{LHV}/kg_{CO2} are reported in the literature. This result shows that adoption of $Ca(OH)_2$ negatively impacts the plant performance due to the energy penalty associated to the storage of the sorbent at nearly ambient temperature. On the other side, this design allows to unlock the possibility to implement a low CAPEX CO_2 capture process, that in turn may result in positive economic indicators, with proper electricity prices. This work will progress in the framework of the BackCap project, focusing on the analysis of the WHR section and on the economic analysis.

Acknowledgments

This project has received funding from the Research Fund for Coal and Steel under grant agreement No 101034000

References

[1] International Energy Agency (IEA), ‘World Energy Outlook 2021’, 2021. [Online]. Available: www.iea.org/weo

[2] E. Grubert and M. Zacarias, ‘Paradigm shifts for environmental assessment of decarbonizing energy systems: Emerging dominance of embodied impacts and design-oriented decision support needs’, *Renewable and Sustainable Energy Reviews*, vol. 159. Elsevier Ltd, May 01, 2022. doi: 10.1016/j.rser.2022.112208.

[3] M. Jafari, A. Botterud, and A. Sakti, ‘Decarbonizing power systems: A critical review of the role of energy storage’, *Renewable and Sustainable Energy Reviews*, vol. 158. Elsevier Ltd, Apr. 01, 2022. doi: 10.1016/j.rser.2022.112077.

[4] B. Arias, Y. A. Criado, and J. C. Abanades, ‘Thermal Integration of a Flexible Calcium Looping CO2 Capture System in an Existing Back-Up Coal Power Plant’, *ACS Omega*, vol. 5, no. 10, pp. 4844–4852, Mar. 2020, doi: 10.1021/ACSOMEGA.9B03552/SUPPL_FILE/AO9B03552_SI_001.PDF.

[5] B. P. Company, ‘Full report – Statistical Review of World Energy 2021’.

[6] M. Astolfi, E. de Lena, and M. C. Romano, ‘Improved flexibility and economics of Calcium Looping power plants by thermochemical energy storage’, *International Journal of Greenhouse Gas Control*, vol. 83, pp. 140–155, Apr. 2019, doi: 10.1016/J.IJGGC.2019.01.023.

[7] M. Astolfi, E. de Lena, F. Casella, and M. C. Romano, ‘Calcium looping for power generation with CO2 capture: The potential of sorbent storage for improved economic performance and flexibility’, *Applied Thermal Engineering*, vol. 194, p. 117048, Jul. 2021, doi: 10.1016/J.APPLTHERMALENG.2021.117048.

[8] Y. A. Criado, B. Arias, and J. C. Abanades, ‘Calcium looping CO2 capture system for back-up power plants’, *Energy & Environmental Science*, vol. 10, no. 9, pp. 1994–2004, Sep. 2017, doi: 10.1039/C7EE01505D.

[9] D. P. Hanak, C. Biliyok, and V. Manovic, ‘Calcium looping with inherent energy storage for decarbonisation of coal-fired power plant’, *Energy & Environmental Science*, vol. 9, no. 3, pp. 971–983, Mar. 2016, doi: 10.1039/C5EE02950C.

[10] C. Ortiz, M. C. Romano, J. M. Valverde, M. Binotti, and R. Chacartegui, ‘Process integration of Calcium-Looping thermochemical energy storage system in concentrating solar power plants’, *Energy*, vol. 155, pp. 535–551, Jul. 2018, doi: 10.1016/J.ENERGY.2018.04.180.

[11] X. Chen, D. Zhang, Y. Wang, X. Ling, and X. Jin, ‘The role of sensible heat in a concentrated solar power plant with thermochemical energy storage’, *Energy Conversion and Management*, vol. 190, pp. 42–53, Jun. 2019, doi: 10.1016/J.ENCONMAN.2019.04.007.

[12] B. Arias, Y. A. Criado, B. Pañeda, and J. C. Abanades, ‘Carbonation Kinetics of Ca(OH)2under Conditions of Entrained Reactors to Capture CO2’, *Industrial and Engineering Chemistry Research*, vol. 61, no. 9, pp. 3272–3277, Mar. 2022, doi: 10.1021/ACS.IECR.1C04888/ASSET/IMAGES/LARGE/IE1C04888_0008.JPEG.

[13] A. A. Scaltsoyiannes and A. A. Lemonidou, ‘On the factors affecting the deactivation of limestone under calcium looping conditions: A new comprehensive model’, *Chemical Engineering Science*, vol. 243, Nov. 2021, doi: 10.1016/J.CES.2021.116797.

[14] R. Han *et al.*, ‘Progress in reducing calcination reaction temperature of Calcium-Looping CO2 capture technology: A critical review’, *Chemical Engineering Journal*, vol. 450, p. 137952, Dec. 2022, doi: 10.1016/J.CEJ.2022.137952.

[15] S. A. Salaudeen, B. Acharya, and A. Dutta, ‘CaO-based CO2 sorbents: A review on screening, enhancement, cyclic stability, regeneration and kinetics modelling’, *Journal of CO2 Utilization*, vol. 23, pp. 179–199, Jan. 2018, doi: 10.1016/J.JCOU.2017.11.012.

[16] ‘BackCap RFCS European Project - CO2 capture from back-up coal power plants using Ca(OH)2’. https://ec.europa.eu/info/funding-tenders/opportunities/portal/screen/how-to-participate/org-details/995543981/project/101034000/program/31061225/details (accessed Aug. 22, 2022).

[17] M. Spinelli, I. Martínez, and M. C. Romano, ‘One-dimensional model of entrained-flow carbonator for CO2 capture in cement kilns by Calcium looping process’, *Chemical Engineering Science*, vol. 191, pp. 100–114, Dec. 2018, doi: 10.1016/J.CES.2018.06.051.

[18] V. Materic and S. I. Smedley, ‘High temperature carbonation of Ca(OH)2’, *Industrial and Engineering Chemistry Research*, vol. 50, no. 10, pp. 5927–5932, May 2011, doi: 10.1021/IE200367W/ASSET/IMAGES/LARGE/IE-2011-00367W_0004.JPEG.

[19] A. Rahman, M. G. Rasul, M. M. K. Khan, and S. Sharma, ‘Aspen Plus Based Simulation for Energy Recovery from Waste to Utilize in Cement Plant Preheater Tower’, *Energy Procedia*, vol. 61, pp. 922–927, Jan. 2014, doi: 10.1016/J.EGYPRO.2014.11.996.

[20] D. Alfani, M. Binotti, E. Macchi, P. Silva, and M. Astolfi, ‘sCO2 power plants for waste heat recovery: design optimization and part-load operation strategies’, *Applied Thermal Engineering*, vol. 195, p. 117013, Aug. 2021, doi: 10.1016/J.APPLTHERMALENG.2021.117013.

[21] B. Hu, J. Guo, Y. Yang, and Y. Shao, ‘Optimization of low temperature geothermal organic Rankine power generation system’, *Energy Reports*, vol. 8, pp. 129–138, Jun. 2022, doi: 10.1016/J.EGYR.2022.01.101.

[22] M. Astolfi, D. Alfani, S. Lasala, and E. Macchi, ‘Comparison between ORC and CO2 power systems for the exploitation of low-medium temperature heat sources’, *Energy*, vol. 161, pp. 1250–1261, Oct. 2018, doi: 10.1016/J.ENERGY.2018.07.099.

[23] ‘European Best Practice Guidelines for Assessment of Carbon Dioxide (CO2) Capture Technologies | Climate Technology Centre & Network | Fri, 09/08/2017’. https://www.ctc-n.org/resources/european-best-practice-guidelines-assessment-carbon-dioxide-co2-capture-technologies (accessed Aug. 22, 2022).

[24] G. Grasa, R. Murillo, M. Alonso, and J. C. Abanades, ‘Application of the random pore model to the carbonation cyclic reaction’, *AIChE Journal*, vol. 55, no. 5, pp. 1246–1255, May 2009, doi: 10.1002/AIC.11746.

[25] Y. A. Criado and B. Arias, ‘Analysis of Operation Conditions of Ca(OH)2 Entrained Carbonator Reactors for CO2 Capture in Backup Power Plants’, *ACS Omega*, vol. 7, no. 32, pp. 28093–28100, Aug. 2022, doi: 10.1021/ACSOMEGA.2C02134.