

Technical and economic feasibility evaluation of calcium looping with no CO₂ recirculation

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ABSTRACT

Carbon capture and storage is expected to play a key role in decarbonisation of the power and industrial sectors, with calcium looping (CaL) being regarded as a well-developed technology that can reduce energy and economic penalties associated with mature technologies. Nevertheless, retrofits of CaL to coal-fired power plants result in net efficiency penalties of 5–8% points. The state-of-the-art CaL configurations assume that, similarly to oxy-fuel combustion systems, CO₂ needs to be recycled to moderate the temperatures in the calciner. This study aims to assess the feasibility of CaL with no CO₂ recirculation to the calciner via pilot plant testing and techno-economic analysis. The results collected during the experimental trials indicated that the temperatures inside of the calciner were within 930–950°C, which were within the commonly reported operating temperature range of that reactor. Furthermore, the techno-economic analysis of the CaL retrofit to a conventional 580 MW_{el} coal-fired power plant indicated that operation of CaL without CO₂ recirculation will have a negligible effect on the net thermal efficiency of the entire system. Nevertheless, reduction in the size of the system resulted in a 21.7% reduction in the specific capital cost, and thus a 14.3% and 27.4% reduction in the levelised cost of electricity and cost of CO₂ avoided, respectively. Therefore, CaL with no CO₂ recirculation can be considered as a technically and economically feasible option to reduce the economic penalty associated with this emerging technology.

Key Words: Calcium looping, pilot plant, clean coal, techno-economic analysis, process modelling, economic penalty reduction

1 INTRODUCTION

To mitigate the severe consequences of climate change, the mean earth temperature rise compared to that of the pre-industrial level should be kept well below 2°C by 2100 (2DS) [1] and efforts to limit it to 1.5°C need to be pursued [2]. Accordingly, it has been targeted to reduce greenhouse gas emissions by at least 80% from the 1990 baseline by 2050, in order to achieve the 2DS targets [3,4]. Fossil-fuel-fired power plants and carbon-intensive industries are considered as the major CO₂ emitters and are obligated to significantly reduce their CO₂ emissions [5,6]. Carbon capture and storage (CCS) is regarded as the key option for decarbonisation of the power and industrial sectors [7], and in 2DS is expected to contribute to nearly 14% reduction in global CO₂ emissions by 2050 [8]. It has been estimated that the exclusion of CCS can cause up to 70% increase in the cost of achieving emission reduction targets [3].

Nevertheless, the CO₂ capture technologies closest to commercialisation, such as amine scrubbing and oxy-fuel combustion, are associated with relatively high efficiency and economic penalties. Thus, the implementation of such systems in the power sector can cause a drastic increase in the cost of electricity, which makes CCS not economically favoured at the moment [9]. Calcium looping (CaL) is a promising CO₂ capture technology that is based on the reversible carbonation reaction of lime and CO₂ at a high temperature [10]. In this process, the Ca-based material initially undergoes the carbonation reaction with CO₂ present in the flue gas from power plants or carbon-intensive industry processes, at 600–700°C. Subsequently, the saturated sorbent is transferred to the calciner and regenerated at 900–950°C [11]. The heat required for regeneration of the solid sorbent is typically provided by direct oxy-fuel combustion in the calciner, to obtain a high-purity CO₂ stream. Importantly, CaL has already been demonstrated at megawatt-scale pilot

plants, such as the 1 MW_{th} unit in Darmstadt [12], the 1.7 MW_{th} unit in Oviedo (La Pereda) [13], and the 1.9 MW_{th} unit at the Industrial Technology Research Institute (ITRI), Taiwan [14].

It has been estimated that the retrofit of CaL to a conventional coal-fired power plant can lead to an efficiency penalty of 5–8% [15–17], which is significantly lower compared to that of amine scrubbing (9.5–12.5%) [18] and oxy-fuel combustion (8–12%) technologies [19]. Importantly, the high-grade heat available in CaL is often utilised to raise high-pressure steam for the secondary steam cycle, leading to increased power output of the entire system by more than 50% [20,21]. In addition, CaL retrofits are characterised by competitive economic performance in terms of levelised cost of electricity and cost of CO₂ avoided (LCOE = 54.3–96 €/MW_{el}h; AC = 28.9–58.3 €/tCO₂) [22–24] compared to chemical solvent scrubbing (LCOE = 65–89 €/MW_{el}h; AC = 20–60 €/tCO₂) [25–27] and oxy-fuel combustion (LCOE = 55–75 €/MW_{el}h; AC = 35–75 €/tCO₂) [28,29] retrofits. One of the main sources of the parasitic load in CaL is associated with the air separation unit that provides high-purity O₂ for oxy-fuel combustion in the calciner [9]. Therefore, further reduction in the efficiency and economic penalties of CaL can be achieved by eliminating the air separation unit and utilising an alternative heat source for sorbent regeneration, such as chemical looping [30,31], as well as indirect heat transfer from an external heat source using heat carriers [32,33], heat pipes [34,35] or heat transfer walls [36,37].

An alternative option to further improve the economic performance of the CaL process may be achieved by reducing, or even eliminating, the CO₂ recirculation back to the calciner to dilute the high-purity O₂ stream fed to the calciner. This strategy can potentially reduce the calciner heat requirement, the size of the calciner and the entire CaL system. Consequently, a decrease in the operating and capital

costs can be expected. In addition, such strategy can reduce the complexity of the CaL system, considering its design and operation. In CaL, the heat generated via oxy-fuel combustion in the calciner in the absence of the recycled CO_2 is moderated through the endothermic calcination reaction and by the sorbent fed to the calciner at lower temperature than the desired operating temperature of that reactor. As a result, the calciner can be perceived as a heat sink. This is opposite to the oxy-fuel combustion systems, where no heat sink is present and CO_2 needs to be recirculated to moderate the temperatures in the oxy-combustors. Such concept has been recently evaluated via computational fluid dynamics modelling of the calciner based on a test run in the La Pereda pilot plant [38]. However, the pilot-plant testing activities were only conducted with O_2 concentrations below 50%_{vol} [15] and CaL with high O_2 concentration in the calciner was only considered in thermodynamic assessments of CaL retrofits to fossil-fuel-fired power plants [39,40]. Therefore, it is still not clear whether operation of the calciner under high O_2 concentration is a feasible solution from technical and economic standpoints. This study aims to assess the technical feasibility of CaL with no CO_2 recirculation by demonstrating the stable operation of a 25 kW_{th} CaL pilot plant. It also aims to present a techno-economic analysis that evaluates the thermodynamic and economic feasibility of the considered concept for decarbonisation of a conventional 580 MW_{el} coal-fired power plant.

2 EXPERIMENTAL SET-UP

The experimental set-up used in this work has been previously described elsewhere [11]. It comprises a circulating fluidised bed carbonator (4.3 m tall, 0.1 m ID) and a bubbling fluidised bed calciner (1.2 m tall, 0.165 m ID), as presented in Figure 1. The solids are transferred from one reactor to the other via two loop seals fluidised with

nitrogen. Both reactors are electrically heated. In addition, there are gas preheaters in the gas lines entering both reactors. Natural gas is directly fed into the fluidised bed of the calciner to produce the heat needed for the calcination reaction via combustion. For the operational mode considered in this work, the start-up procedure consisted of increasing the O_2 concentration up to 100%_{vol} in the gas stream fed to the calciner in successive 20%_{vol} steps from the initial value of 40%_{vol}. This aimed to ensure stable combustion in the calciner while the O_2 concentration was being increased. This was monitored by continuous analysis of the gas composition at the outlet of the calciner and the calciner temperature. In this experiment, Longcal limestone was used as a Ca-based sorbent with a particle size range of 200–300 μm . The superficial gas velocities were maintained at 2.5 m/s in the carbonator and 0.3 m/s in the calciner.

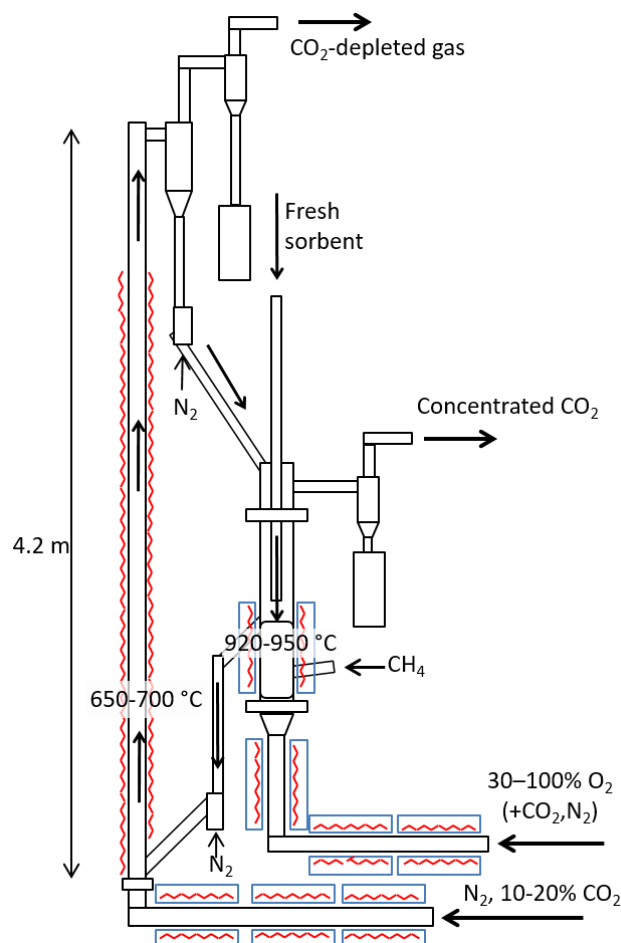


Figure 1: Experimental set-up of the pilot plant

3 TECHNO-ECONOMIC MODEL DESCRIPTION

3.1 Supercritical coal-fired power plant

A supercritical coal-fired power plant used as a reference in this study is characterised with a gross power output of 580 MW_{el} and a net thermal efficiency of 38.0%_{HHV} [41,42]. A process model of this unit, for which the key assumptions are summarised in Table 1, has been previously developed in Aspen Plus® [41,42] based on, and validated with data presented in, the revised NETL report [43].

The power plant considered in this study comprises a power boiler, with NO_x and SO_x emission control equipment, as well as an electrostatic precipitator. In the power boiler, heat from the combustion of coal is utilised to raise high-pressure steam for the conventional steam cycle. The steam generator comprises primary, secondary, and reheat superheaters, as well as an economiser, in which the temperature of the live (242.3 bar) and reheat steam (45.2 bar) is raised to 593.3°C. These are fed to the steam turbine section that comprises high- (HP), intermediate- (IP) and low-pressure (LP) extraction condensing steam turbines. The design isentropic efficiencies of the steam turbine sections were determined to match the steam discharge conditions [43]. To improve the steam cycle efficiency, a part of the steam from the turbine sections is extracted to feed the feedwater heating train. This section consists of five LP feedwater heaters, the last of which is a mixed feedwater heater called a deaerator, and three HP feedwater heaters.

3.2 Calcium looping post-combustion CO₂ capture plant

A process model for calcium looping, which is considered as a post-combustion CO₂ capture unit in this study, was developed in Aspen Plus[®] by Hanak et al. [9] and found to be in good agreement with the experimental data at different flue gas loads.

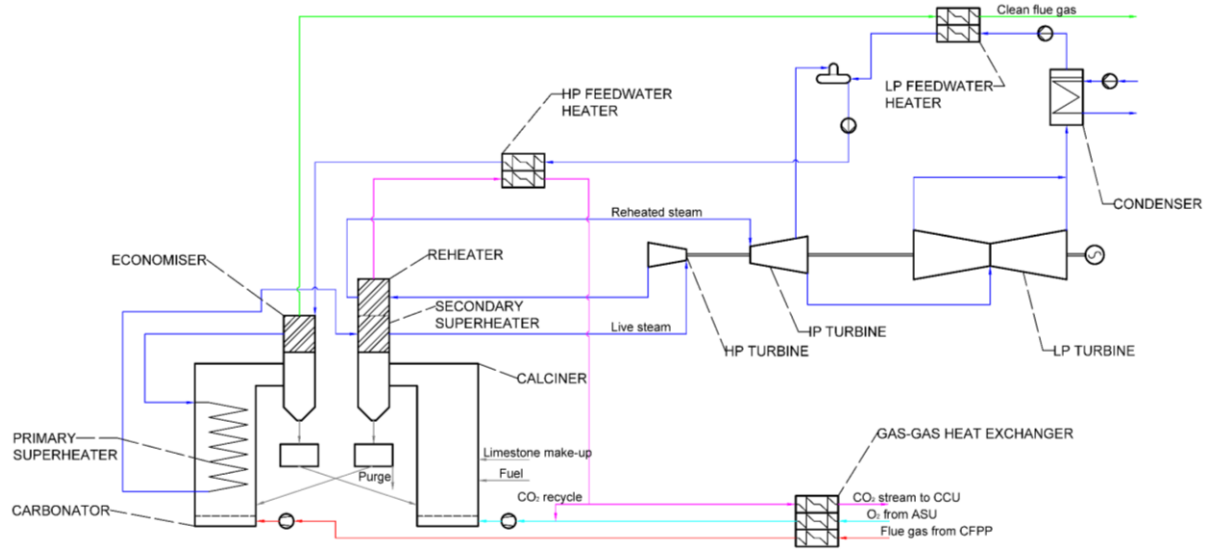


Figure 2: Process flow diagram of the calcium looping process for decarbonisation of a conventional 580 MW_{el} coal-fired power plant

Calcium looping (Figure 2) comprises a carbonator, which is modelled as a stoichiometric reactor, and a calciner, which is modelled as a Gibbs reactor. To sustain the endothermic calcination reaction and ensure that high-purity CO₂ leaves the calciner, an additional amount of coal is oxy-combusted directly in this reactor. The coal decomposition is modelled using a yield reactor. Moreover, the standard double-column cryogenic air separation unit, the model of which was described in detail by Hanak et al. [9], is considered as the source of high-purity O₂ (95%_{vol}). The average sorbent conversion in the carbonator is estimated using the semi-empirical correlation derived by Rodríguez et al. [44], which is shown in Eq. (1). In this correlation, the maximum average conversion of sorbent is a function of the carbonation (f_{carb}) and calcination extent (f_{calc}), sorbent characteristics (a_1 , a_2 , f_1 , f_2 ,

b), fraction of never-calcined limestone in the system (r_0), fresh limestone make-up (F_0) and solid looping rate (F_R). The sorbent deactivation curve derived from experimental tests at the 1.7 MW_{th} INCAR pilot plant [45] is used. Moreover, to account for equilibrium limitations, the equilibrium partial pressures of CO₂ in the gas streams leaving the carbonator and the calciner are determined using a correlation by Baker [46].

$$X_{ave} = (F_0 + F_R r_0) f_{calc} \left[\frac{a_1 f_1^2}{F_0 + F_R f_{carb} f_{calc} (1 - f_1)} + \frac{a_2 f_2^2}{F_0 + F_R f_{carb} f_{calc} (1 - f_2)} + \frac{b}{F_0} \right] \quad (1)$$

Table 1: Summary of the key process model assumptions

Parameter		Value
Coal-fired power plant		
Combustor	Excess air ratio (% _{vol})	20.0
Primary steam cycle	Design live/reheat steam temperature (°C)	593.3/593.3
	Design live/reheat steam pressure (bar)	242.3/45.2
	Final feedwater temperature (°C)	291.0
	Feedwater heater terminal temperature difference (°C)	2.8
	Feedwater heater minimum temperature approach (°C)	5.6
	Isentropic efficiency of compressors (%)	80.0
	Isentropic efficiency of pumps (%)	80.0
	Electrical efficiency of generator (%)	97.5
	SO ₂ removal efficiency (%)	98.0
Flue gas treatment	Fly ash removal efficiency (%)	99.8
	NO _x removal efficiency (%)	86.0
Calcium looping CO₂ capture plant		
Carbonator	Temperature (°C)	650
	Carbonation extent (-)	0.70
Calciner	Temperature (°C)	900
	Calcination extent (-)	0.95
	Relative make-up (fresh limestone/sorbent circulation rate) (-)	0.04
	O ₂ excess (% _{vol,dry})	2.5
Secondary steam cycle	Design live/reheat steam temperature (°C)	593.3/593.3
	Design live/reheat steam pressure (bar)	242.3/45.2
	Isentropic efficiency of high-pressure steam turbine (%)	92.0
	Isentropic efficiency of intermediate-pressure steam turbine (%)	94.0
Auxiliary equipment	Isentropic efficiency of low-pressure steam turbine (%)	88.0

CO ₂ compression unit	Polytropic efficiency of CO ₂ compressors (%)	78.0–80.0
	Isentropic efficiency of CO ₂ pump (%)	80.0
	Intercooling temperature (°C)	40.0
	CO ₂ final pressure (bar)	110.0
	CO ₂ final temperature (°C)	30.0
	O ₂ purity (% _{vol,dry})	95.0
Air separation unit	Polytropic efficiency of air compressors (%)	79.0–80.0
	Intercooling temperature (°C)	40.0
	Final air pressure (bar)	5.8

High-grade heat available in the carbonator and in the process streams is utilised to raise high-pressure steam for the secondary steam cycle, which is characterised with similar operating conditions to those in the conventional 580 MW_{el} coal-fired power plant (Table 1).

Finally, it is established that the CO₂ stream pressure at ambient temperature required for pipeline transport is 110 bar [47]. Therefore, the nearly pure CO₂ stream from the CaL process is first compressed to 80 bar and then pumped to 110 bar prior to transport. Due to equipment limitations, the CO₂ compressor is modelled as a set of nine compression stages with constant stage polytropic efficiency (78–80%) [48,49]. Each compression stage consists of a centrifugal compressor, stage intercoolers and scrubbers. The CO₂ pump is assumed to have an isentropic efficiency of 80%.

3.3 Techno-economic analysis methodology

In this study, the techno-economic performance of the CaL retrofit is evaluated in two scenarios: the first considers CO₂ recirculation level of 52.8% to ensure O₂ concentration in the gas stream entering the calciner is maintained at 30%_{vol} (with recirculation); the second considers no CO₂ recirculation (without recirculation) and assumes that a high-purity O₂ stream is directly fed from the air separation unit to the

calcliner without dilution with CO₂ recycle gas. The performance is then benchmarked against that of the conventional 580 MW_{el} coal-fired power plant. Moreover, a parametric study on the CO₂ recirculation level is conducted to vary the O₂ concentration in the gas stream entering the calcliner between 30%_{vol} and 95%_{vol}.

The thermodynamic performance of the considered scenarios is characterised using the key performance indicators that are commonly used to assess the performance of conventional power generation systems. Primarily, these are the net power output ($\dot{W}_{net}$) and net thermal efficiency (η_{th}), which is defined in Eq. (2) as the ratio of the net power output and the heat input from fuel combustion ($\dot{Q}_{fuel}$). In addition, the net efficiency penalty (EP) defined in Eq. (3) is calculated to benchmark the performance of the CaL retrofits against the 580 MW_{el} coal-fired power plant. Finally, environmental performance is represented in Eq. (4) as the specific CO₂ emission (e_{CO_2}) defined as the ratio of CO₂ emission rate ($\dot{m}_{CO_2}$) and the net power output.

$$\eta_{th} = \frac{\dot{W}_{net}}{\dot{Q}_{fuel}} \quad (2)$$

$$EP = \eta_{th,ref} - \eta_{th,CaLC} \quad (3)$$

$$e_{CO_2} = \frac{\dot{m}_{CO_2}}{\dot{W}_{net}} \quad (4)$$

The economic performance of the calcium looping retrofit to the conventional 580 MW_{el} coal-fired power plant is assessed in terms of the levelised cost of electricity and cost of CO₂ avoided that are estimated using Eq. (5) and Eq. (6), respectively [24,50,51].

$$LCOE = \frac{TCR \times FCF + FOM}{\dot{W}_{net} \times CF \times 8760} + VOM + \frac{SFC}{\eta_{th}} \quad (5)$$

$$AC = \frac{LCOE_{capture} - LCOE_{ref}}{e_{CO_2,ref} - e_{CO_2,capture}} \quad (6)$$

These parameters correlate the thermodynamic performance indicators of the retrofitted system, such as net power output, net thermal efficiency, capacity factor (CF), and specific CO_2 emissions, with its economic performance indicators, such as total capital requirement (TCR), variable (VOM) and fixed (FOM) operating and maintenance costs, specific fuel cost (SFC), and the fixed charge factor (FCF), which considers the system's lifetime and project discount rate. In this analysis, the capital cost of the reference coal-fired power plant is determined using the exponential method function [52]. Conversely, the capital cost of the calcium looping process (C_{CaL}) is estimated using the correlation developed by Romano et al. [23] and represented by Eq. (7). This approach considers the capital cost for an oxy-fuel circulating fluidised-bed system as the reference capital cost (C_0) [53], as well as allows taking into account the reactor volume (V), the heat input in the calciner ($\dot{Q}_{calc}$), along with the corresponding scaling factors, and the fraction of the total cost of the calciner associated with the heat transfer surfaces (α).

$$C_{CaL} = C_0 \left[\alpha \left(\frac{\dot{Q}_{calc}}{\dot{Q}_{0,calc}} \right)^{SF,Q} + (1 - \alpha) \left(\frac{V_{calc}}{V_{0,calc}} \right)^{SF,V} + (1 - \alpha) \left(\frac{V_{carb}}{V_{0,carb}} \right)^{SF,V} \right] \quad (7)$$

The fixed and variable operating and maintenance costs are expressed as a fraction of total capital cost, while operating costs associated with fuel and sorbent consumption, CO_2 storage and transport, and CO_2 emission are determined based on process simulation outputs and specific costs of these components using economic data from Table 2.

Table 2: Economic model assumptions

Parameter		Value
Reference coal-fired power plant	Reference equipment specific capital cost (€/kW _{el}) [54,55]	1100.0
Calcium looping combustion process	Reference equipment specific capital cost (€/kW _{el}) [53]	1252.3
Scaling factor	Reference coal-fired power plant (-) [53]	0.67
	Reactor volume (-) [23]	0.67
	Heat input to the combustor (-) [23]	0.90

Other economic parameters	Fraction of total system cost associated with heat transfer surfaces (-) [23]	0.85
	Variable cost as a fraction of total capital cost (%) [24,56]	2.0
	Fixed cost as a fraction of total capital cost (%) [24,56]	1.0
	Carbon tax (€/tCO ₂) [24,56]	0.0
	Raw sorbent cost (€/t) [24,56]	6.0
	CO ₂ transport and storage cost (€/tCO ₂) [57]	7.0
	Coal price (€/t) [56,58]	40.5
	Expected lifetime (years) [24,56]	25
	Project interest rate (%) [24,56]	8.78
	Capacity factor (%) [24,56]	80

4 RESULTS AND DISCUSSION

4.1 Experimental demonstration

The results collected during the experimental trials indicated that it is feasible to operate the calciner under high O₂ concentration, as the temperatures in the middle and top section of the reactor were at acceptable levels (Figure 3). It can be seen that the temperature was maintained relatively constant between 930 and 950°C during steady state operation with no CO₂ recirculation and 100%_{vol} O₂ concentration in the gas stream entering the calciner. This proves the feasibility of such operational mode at the pilot scale with natural gas burned in the calciner. A uniform temperature distribution and lack of hot-spots in the calciner under high O₂ concentration operation can be primarily associated to both the endothermic nature of the calcination reaction and the requirement for preheating of the sorbent entering that reactor. This is because both factors can be perceived as heat sinks that moderate the temperature increase due to the heat generation from oxy-fuel combustion under high O₂ concentration in the calciner. This is opposite to the oxy-fuel combustion systems, where no heat sink is present and CO₂ needs to be recirculated to moderate the temperatures in the oxy-combustors.

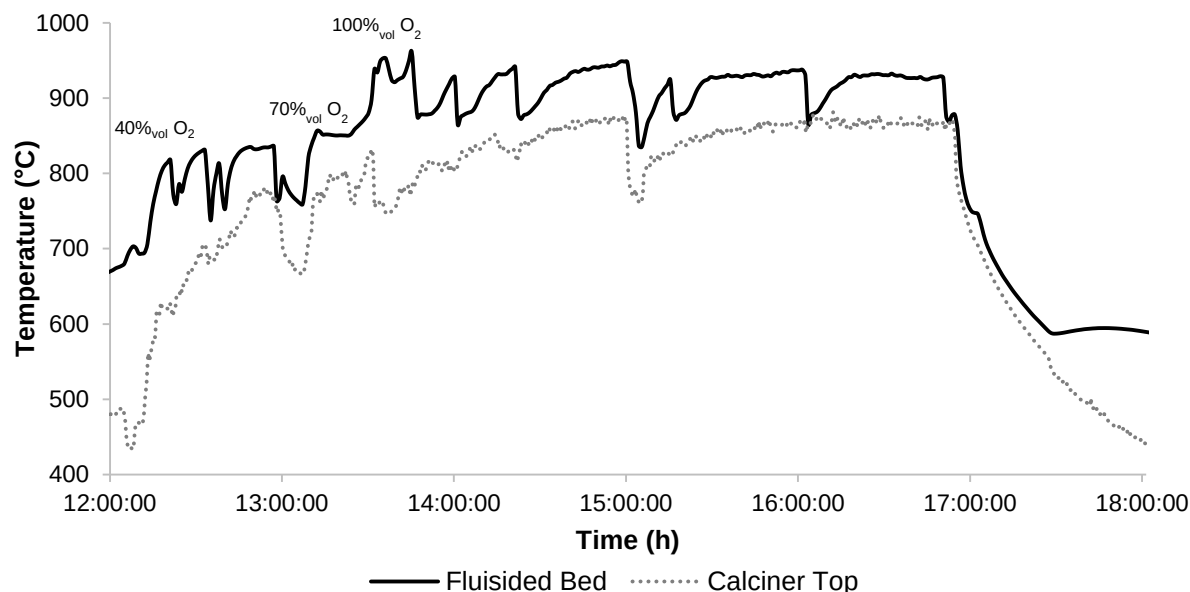
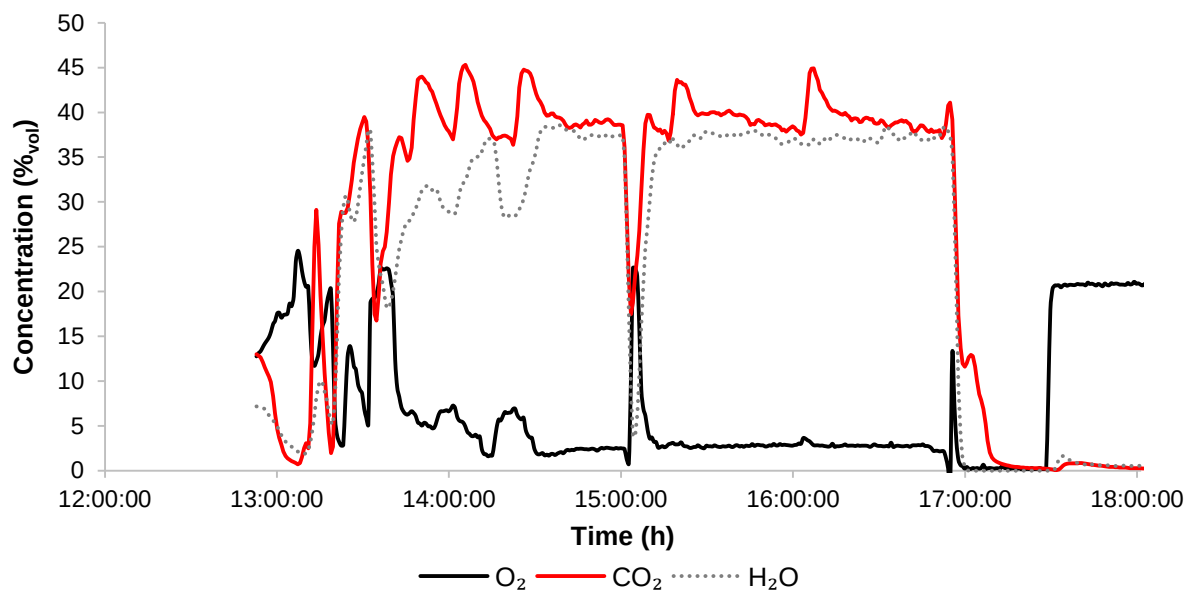
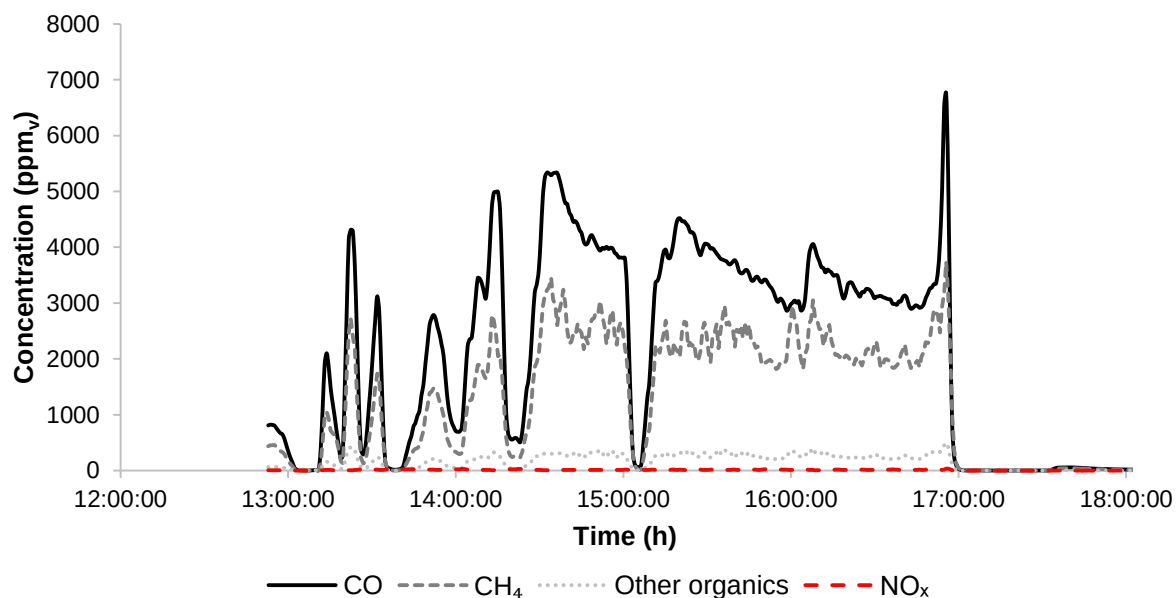


Figure 3: Temperature in the calciner throughout the test

Importantly, lack of CO₂ recirculation to moderate the temperature in the calciner has a negligible effect on the combustion performance in that reactor (Figure 4). Namely, the outlet gas comprised mainly CO₂ and steam, with some impurities containing mainly O₂, CO and CH₄. These results were similar to those with CO₂ recirculation, but the main difference was the amount of steam present in the gas leaving the calciner. Also, operation under high O₂ concentration was found not to result in increased NO_x generation in the calciner. Therefore, this operational mode is feasible and has several advantages including less complex design. Also, the CaL system designed accordingly is expected to be easier to operate.



a)



b)

Figure 4: Gas composition of a) major and b) trace species at the outlet of the calciner

4.2 Techno-economic performance evaluation

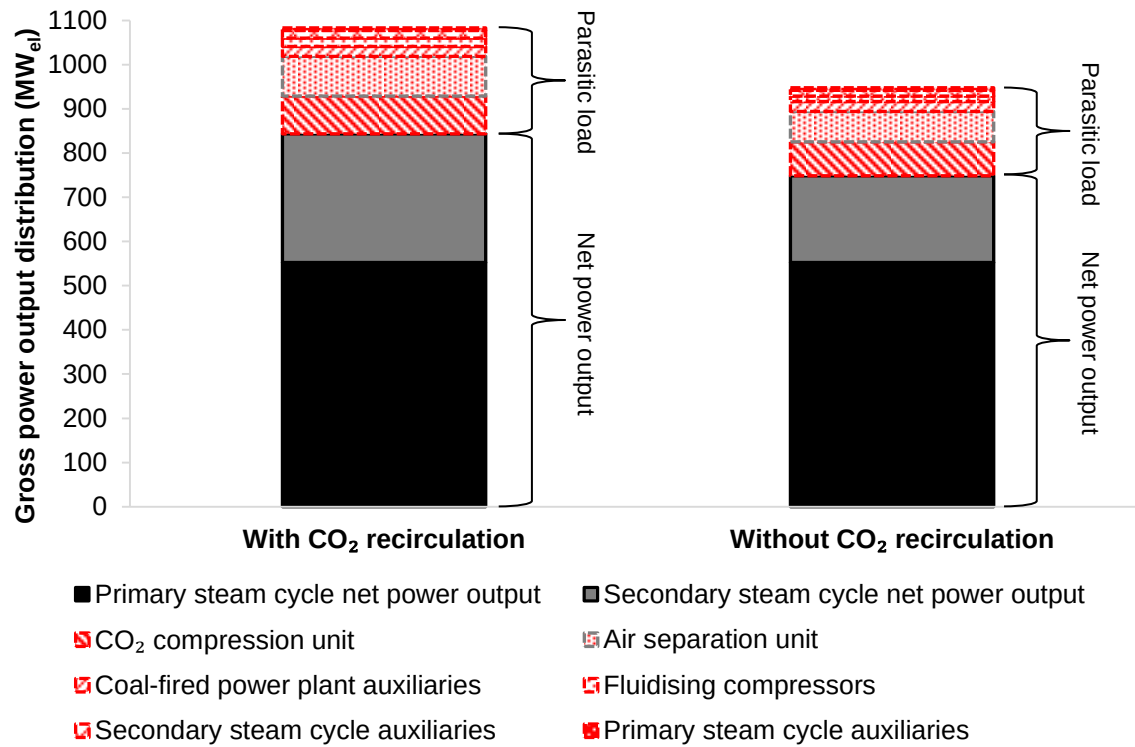
4.2.1 Proof of concept

Following the successful pilot plant demonstration of CaL operation with no CO₂ recirculation, the techno-economic analysis was performed to assess the potential benefits of such operating mode at a large scale (Table 3). The analysis of the thermodynamic performance indicated that operating CaL without CO₂ recirculation

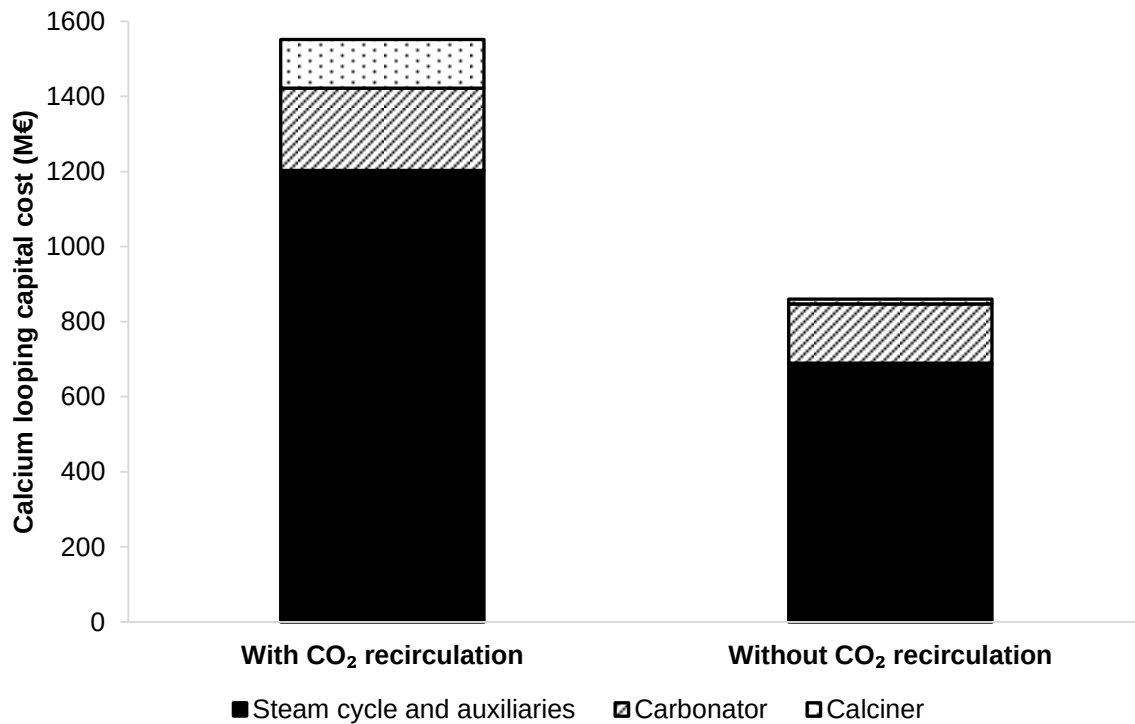
results in an 11.2% reduction in the net power output of the entire system, compared to CaL with CO₂ recirculation. This can be explained by a drop in the total coal consumption rate from 102.4 kg/s (with CO₂ recirculation) to 90.9 kg/s (without CO₂ recirculation). This, in turn, can be related to a 23.6% drop in the heat requirement in the calciner, primarily due to no need for preheating the CO₂ recycle stream to the required calciner operating temperature. It is important to highlight that the magnitude of the reduction in both the net power output of the entire system and the total coal consumption was found to be of the same level. Therefore, the operation of CaL without CO₂ recirculation does not have a significant effect on the net efficiency, and thus on the net efficiency penalty associated with such CaL retrofit scenario, as well as specific coal consumption. Therefore, both CaL retrofit scenarios are characterised with a net efficiency penalty of 7.5%_{HHV} points. Such figure is lower compared to amine scrubbing (9.5–12.5%) [18] and oxy-fuel combustion (8–12%) technologies [19]. Yet, in contrast to the retrofits of these technologies, a CaL retrofit results in 52.5% and 35.4% increase in the net power output of the entire system in the scenario with and without CO₂ recirculation, respectively. Such increase in the net power output is in line with other studies assessing retrofits of CaL to fossil-fuel-fired power plants [15,16]. Finally, the thermodynamic performance analysis revealed that to maintain the desired overall CO₂ capture level of 90% in the CaL retrofit scenario without CO₂ recirculation, the CO₂ capture level in the carbonator must be increased by 2.1% points. This is directly related to the reduction in the coal consumption, and thus less CO₂ released from coal oxy-combustion in the calciner. This, along with a reduction in the net power output, resulted in a 1.4% increase in the specific CO₂ emission in the CaL retrofit scenario without CO₂ recirculation.

Table 3: Techno-economic performance indicators

Parameter	Conventional coal-fired power plant	Calcium looping retrofit	
		With CO ₂ recirculation	Without CO ₂ recirculation
Thermodynamic performance indicators			
Gross power output (MW _{el})	580.4	1083.2	947.7
Coal-fired power plant auxiliaries power requirement (MW _{el})	22.0	22.0	22.0
Primary steam cycle auxiliaries power requirement (MW _{el})	5.7	5.7	5.7
Secondary steam cycle auxiliaries power requirement (MW _{el})	-	17.5	12.8
CO ₂ compression unit power requirement (MW _{el})	-	85.5	76.4
Air separation unit power requirement (MW _{el})	-	90.2	69.4
Fluidising compressors power requirement (MW _{el})	-	19.2	12.8
Net power output (MW _{el})	552.7	843.1	748.6
Net thermal efficiency (% _{HHV})	38.0	30.5	30.5
Net efficiency penalty (% _{HHV} points)	-	7.5	7.5
Coal consumption rate (kg/s)	53.8	102.4	90.9
O ₂ requirement (kg/s)	-	109.3	84.1
CO ₂ emission rate (kg/s)	121.6	25.9	23.3
CO ₂ capture rate in carbonator (kg/s)	-	95.7	98.3
Overall CO ₂ capture rate (kg/s)	-	233.6	210.1
Specific coal consumption (g/kW _{el} h)	350.3	437.3	437.3
Specific CO ₂ emission (g/kW _{el} h)	792.3	110.6	112.1
CO ₂ capture level in carbonator (%)	-	78.7	80.8
Overall CO ₂ capture level (%)	-	90.0	90.0
Economic performance indicators			
Part of capital cost associated with calcium looping (%)	-	70.7	57.3
Specific capital cost (€/kW _{el,gross})	1222.6	2026.1	1586.1
Levelised cost of electricity (€/MW _{el} h)	38.0	74.8	64.1
CO ₂ avoided cost (€/tCO ₂)	-	57.3	41.6



a)



b)

Figure 5: Distribution of a) the gross power output and b) calcium looping capital cost

Regardless of reduction in the net power output of the entire system, which would have implications on the revenue from electricity sales in the long term, operation of CaL without CO₂ recirculation will result in a substantial reduction in the footprint of

the CaL plant. This is because the equipment size is expected to be significantly smaller, compared to the scenario with CO₂ recirculation. This can be clearly seen in Figure 5, which compares the distribution of both the gross power output (Figure 5a) and the calcium looping capital cost (Figure 5b) in the considered scenarios. It can be noted that the power requirement of the air separation unit and CO₂ compression unit were 23.0% and 10.7% lower in the scenario without CO₂ recirculation, respectively. In addition, the gross power output of the secondary steam cycle was shown to be 26.9% lower in this scenario. As a result, the capital cost of the secondary steam cycle and the auxiliary units, including air separation and CO₂ compression units, was reduced by 42.7%. Furthermore, an 86% reduction in the volumetric flow rate of gas entering the calciner was achieved in the scenario without CO₂ recirculation. As a result, the capital cost of the calciner was reduced by nearly 90% (Figure 5b). Overall, a 44.5% reduction in the capital cost of CaL was observed in that scenario and this was reflected in a 21.7% reduction in the specific capital cost of the entire system (Table 3). It needs to be noted that such reduction is not only a result of a reduction in the size of CaL equipment, but also a reduction in the size of the heat exchangers after the calciner. Importantly, the reduction in the size significantly improves the economic performance of the CaL retrofit scenario, as the levelised cost of electricity and cost of CO₂ avoided were estimated to be 14.3% and 27.4% lower in the scenario without CO₂ recirculation compared to the scenario with CO₂ recirculation. Considering the fact that CaL without CO₂ recirculation can increase the net power output of the entire system by 35.4% compared to the conventional coal-fired power plant, its economic performance is competitive to that of chemical solvent scrubbing (LCOE = 65–89 €/MW_{el}h; AC = 20–60 €/tCO₂) [25–27]

and oxy-fuel combustion ($LCOE = 55\text{--}75 \text{ €/MW}_{el}h$; $AC = 35\text{--}75 \text{ €/tCO}_2$) [28,29] retrofits.

A further reduction in the parasitic load in the CaL retrofit scenarios can be achieved by reducing the power requirement of the air separation unit and CO_2 compression unit, which account for more than 70% of the total parasitic load of the entire system. This can be achieved by increasing the degree of heat integration in the entire system and utilising alternative options to provide heat for sorbent regeneration, such as chemical looping [30,31,59,60]. Moreover, the combination with other configurations, such as cyclonic preheaters [61] or mixing seal valves [56,62], could lead to additional reductions in the fuel and O_2 requirements, as well as in the size of the CaL equipment. Alternatively, a power cycle with a higher thermal efficiency than that of the supercritical steam cycle, such as the supercritical CO_2 cycle [63], can be implemented to utilise the high-grade heat available in CaL.

4.2.2 Parametric study

Having proven the techno-economic feasibility of the scenario in which CaL is operated without CO_2 recycle, it is pertinent to assess the techno-economic performance of the system operating under different CO_2 recirculation levels. The parametric analysis has revealed that a main drop in the heat input to the calciner (Figure 6a) of 18.6% occurred on a reduction in the CO_2 recirculation level from 52.8% to 21.0%, which correspond to O_2 concentrations of 30%_{vol} and 60%_{vol}. As indicated above, if CaL is operated without CO_2 recirculation, the heat input to the calciner can be reduced by up to 23.6%. A reduction in the net power output of the entire system (Figure 6b) was shown to follow a similar trend. Namely, a drop of up to 8.9% was obtained for a reduction in the CO_2 recirculation level from 52.8% to 21.0%, while operation without CO_2 recirculation resulted in an overall reduction in

the net power output of 11.2%. This implies that the largest reduction in both the size of CaL and the heat input to the calciner would occur on an increase of the O_2 concentration from 30%_{vol} to 60%_{vol}, corresponding to a reduction in the CO_2 recirculation level from 52.8% to 21.0%. Beyond this point, the reduction in both the size of CaL and the heat input to the calciner were still observed, but their magnitude was smaller.

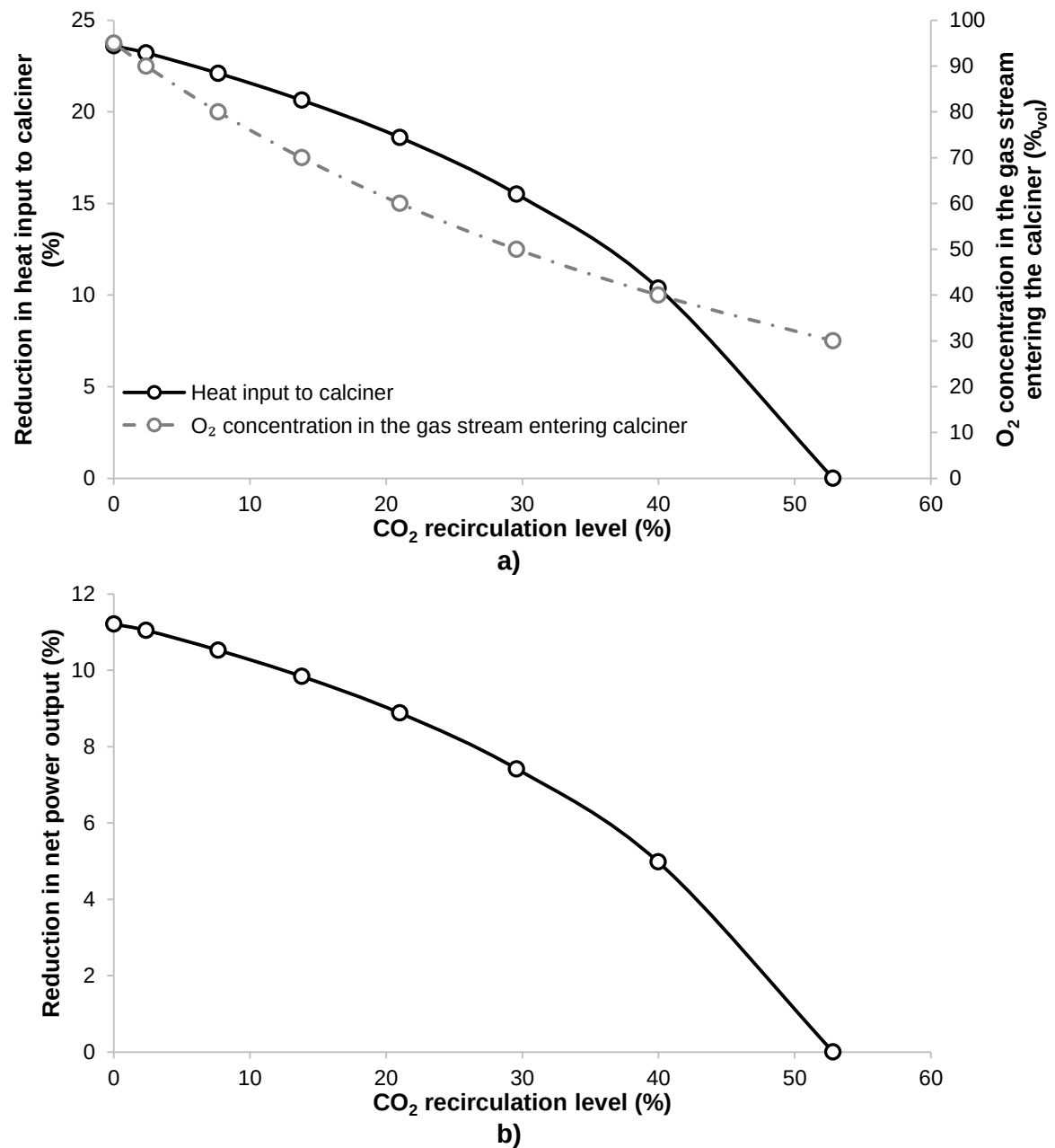


Figure 6: Effect of O_2 concentration in the gas stream entering the calciner on thermodynamic performance of CaL retrofit

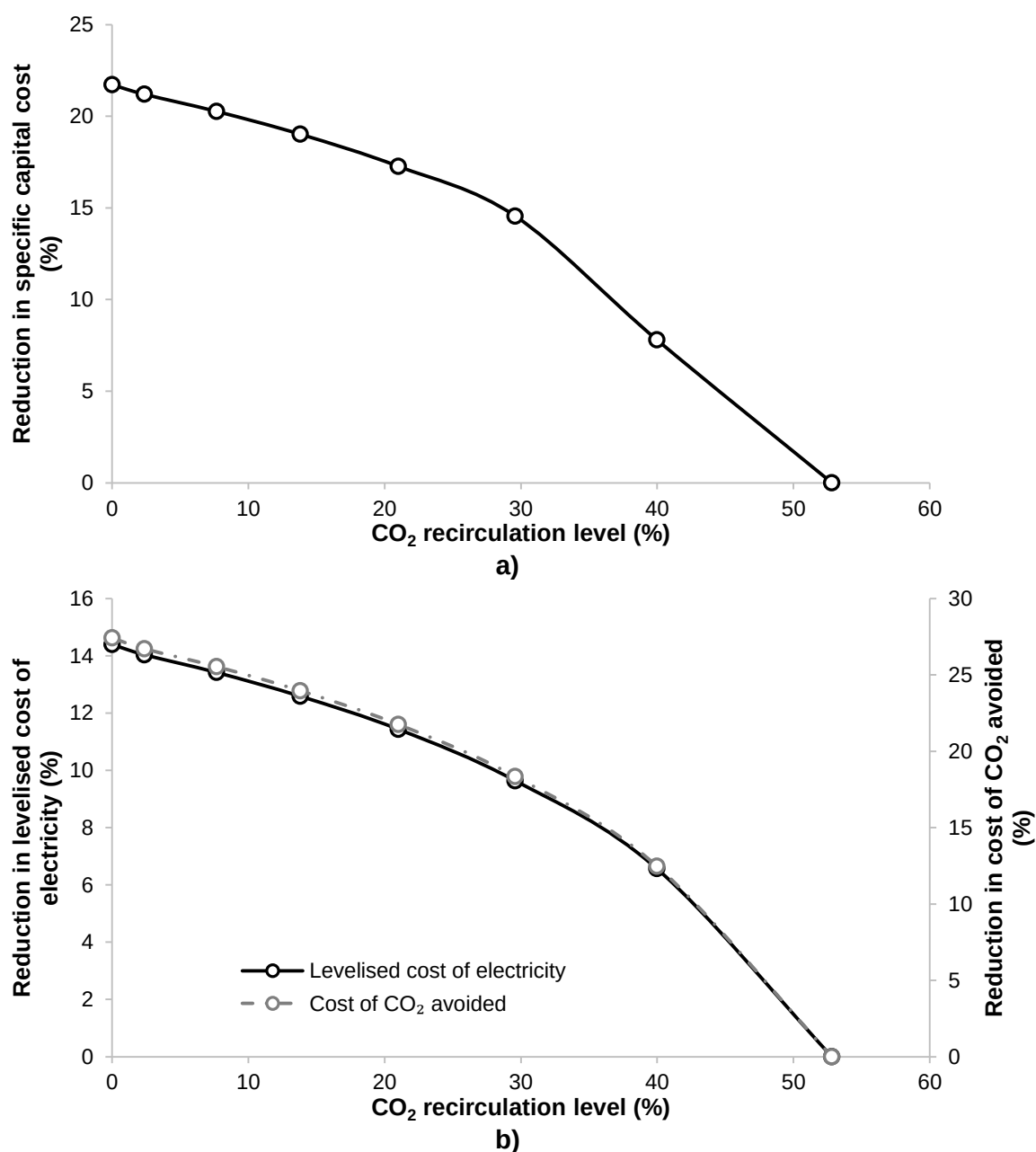


Figure 7: Effect of O₂ concentration in the gas stream entering the calciner on economic performance of CaL retrofit

Figure 7a reveals that a 14.6% reduction in the specific capital cost of the entire system occurs on a reduction in the CO₂ recirculation level from 52.8% to 29.6%. Importantly, if CaL is operated without CO₂ recirculation, the specific capital cost of the entire system is reduced by 21.7%. This can be associated with the observed trend for a reduction in the heat input to the calciner and a higher level of CO₂

capture in the carbonator required to achieve an overall CO₂ capture level of 90%. Such a reduction in the specific capital cost, combined with a higher reduction in the heat input to the calciner than the reduction in the net power output of the entire system, resulted in substantial reductions in the levelised cost of electricity and the cost of CO₂ avoided. Namely, these parameters can be reduced by 9.6% and 18.3% on a reduction in the CO₂ recirculation rate from 52.8% to 29.6%, respectively. If CaL is operated without CO₂ recirculation, these were reduced by 14.4% and 27.4%, respectively.

Importantly, the parametric study has revealed that the largest gain in the economic performance was observed when the CO₂ recirculation rate was reduced from 52.8% to 29.6%. Nevertheless, the operation of CaL without CO₂ recycle has the potential to significantly improve the economic performance of the CaL retrofit scenario and reduce its footprint without having a substantial effect on the thermodynamic performance.

5 CONCLUSIONS

This study aimed to assess the feasibility of CaL with no CO₂ recirculation to the calciner in order to improve the economic performance of a CaL retrofit to a fossil fuel power plant. The technical feasibility of such concept was first tested using the 25 kW_{th} CaL pilot plant. The results collected during the experimental trials indicated that the temperatures inside the calciner were within 930–950°C, which are acceptable from the sorbent degradation standpoint. Moreover, it has been shown that lack of CO₂ recirculation to moderate the temperature in the calciner has a negligible effect on the combustion performance in that reactor. Namely, the concentration of O₂, CO, CH₄ and NO_x were shown to be negligible.

Furthermore, the techno-economic analysis of the CaL retrofit to a conventional 580 MW_{el} coal-fired power plant indicated that operation of CaL without CO₂ recirculation will have a negligible effect on the net thermal efficiency of the entire system. This is because, compared to the CaL with CO₂ recirculation, the reduction in the net power output of the entire system of 11.2% was balanced by 23.6% reduction in the heat requirement in the calciner, which was mainly due to no need for preheating the CO₂ recycle stream to the calciner operating temperature. Importantly, operation of CaL without CO₂ recycle will result in a substantial reduction in the size of the system.

The economic analysis indicated that the retrofit of CaL without CO₂ recirculation will result in 14.3% and 27.4% reduction in the levelised cost of electricity and cost of CO₂ avoided, respectively, compared to CaL with CO₂ recirculation level of 52.8%. This is primary the result of a 21.7% reduction in the specific capital cost. Therefore, CaL with no CO₂ recirculation can be considered as a technically and economically feasible option to reduce the economic penalty associated with this emerging technology. Importantly, in contrast to the mature technologies, CaL without CO₂ recirculation can increase the net power output of the entire system by 35.4% compared to the conventional coal-fired power plant, and its economic performance (LCOE = 64.1 €/MW_{el}h; AC = 41.6 €/tCO₂) is competitive to that of chemical solvent scrubbing (LCOE = 65–89 €/MW_{el}h; AC = 20–60 €/tCO₂) and oxy-fuel combustion (LCOE = 55–75 €/MW_{el}h; AC = 35–75 €/tCO₂).

The evaluated concept of CaL with no CO₂ recirculation to the calciner needs to be further tested, especially for different solid inventories in the calciner and for a broad range of fuels, including coal and biomass. Moreover, further optimisation of the process configuration to improve both the thermodynamic and economic

performance of the retrofitted system should be undertaken. This can be achieved by:

- increasing the level of heat integration in CaL to minimise the heat requirement in the calciner;
- considering alternative O₂ sources for oxy-fuel combustion in the calciner or alternative heat sources for sorbent regeneration to minimise the power requirement associated with O₂ production; and
- utilising the high-grade heat from CaL in a secondary power cycle of higher efficiency.

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NOMENCLATURE

a_1	Sorbent maximum average conversion model fitting parameter	-
a_2	Sorbent maximum average conversion model fitting parameter	-
AC	Cost of CO ₂ avoided	€/tCO ₂
b	Sorbent maximum average conversion model fitting parameter	-
C_0	Reference capital cost of calcium looping process	€/kW _{el}
C_{CaL}	Actual capital cost of calcium looping process	€/kW _{el}
CF	Capacity factor	-
e_{CO_2}	Specific CO ₂ emission	gCO ₂ /kW _{el} h
EP	Net efficiency penalty	% _{HHV} points
F_0	Fresh-limestone make up rate	kmol/s
f_1	Sorbent maximum average conversion model fitting parameter	-
f_2	Sorbent maximum average conversion model fitting parameter	-
f_{calc}	Calcination reaction extent	-
f_{carb}	Carbonation reaction extent	-
FCF	Fixed charge factor	-
FOM	Fixed operating and maintenance cost	€
F_R	CaO looping rate	kmol/s
$LCOE$	Levelised cost of electricity	€/MW _{el} h
$\dot{m}_{CO_2}$	Rate of CO ₂ emission	kg/s
SCF	Specific fuel cost	€/MW _{ch} h
SF, Q	Scaling factor for reactor heat input	-
SF, V	Scaling factor for reactor volume	-
TCR	Total capital requirement	€
$\dot{Q}_{0,calc}$	Reference heat input to the calciner	MW _{th}
$\dot{Q}_{calc}$	Heat input to the calciner	MW _{th}
r_0	Fraction of never-calcined limestone in the system	-
V	Reactor volume	m ³
V_0	Reference volume of reactor	m ³
VOM	Variable operating and maintenance cost	€/MW _{el} h
$\dot{W}_{net}$	Net power output of the retrofitted system	MW _{el}
X_{ave}	Average sorbent conversion	-
α	Fraction of the total cost of a circulating fluidised bed reactor associated with the heat transfer surfaces	-
η_{th}	Net thermal efficiency	-

ABBREVIATIONS

2DS	2°C scenario
CaL	Calcium looping
CCS	Carbon capture and storage
HP	High-pressure
IP	Intermediate-pressure
LP	Low-pressure



DSPACE

<https://dspace.org/>

Technical and economic feasibility evaluation of calcium looping with no CO₂ recirculation

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