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Economic assessment of the swing adsorption reactor cluster for CO_2 capture from cement production

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ABSTRACT

Cement production is responsible for about 8% of global CO_2 emissions. Most of these emissions originate from the process itself and thus cannot be avoided via clean energy, leaving CO_2 capture as the only viable solution. This study investigates the prospects of decarbonizing the cement industry via the swing adsorption reactor cluster (SARC) – a new post-combustion CO_2 capture technology that requires no integration with the host process, consumes only electrical energy and shows a competitive energy penalty. SARC operates by synergistically combining a temperature swing using a heat pump and a vacuum swing using a vacuum pump. In the present study, the SARC concept is evaluated economically and compared to several benchmarks. SARC achieves CO_2 avoidance costs of €52/ton in the base case, which is higher than oxyfuel combustion, similar to calcium looping and lower than four other technology options. SARC can approach the cost of oxyfuel combustion with more optimistic assumptions regarding economies of scale, particularly for the vacuum pump. The local electricity mix is another important factor because SARC, as an electricity consumer, becomes more attractive when the price and CO_2 intensity of electricity is low. Furthermore, the simplicity of retrofitting existing cement plants with the SARC process becomes increasingly valuable when rapid CO_2 emissions reductions are targeted. SARC is therefore well positioned for a global decarbonization effort aiming to limit global warming well below 2 °C.

1. Introduction

Limiting global temperature rise to 1.5 °C will require CO_2 emissions to peak and start a rapid decline within the next decade (IPCC, 2018). Although the stated policies of world nations are not aligned with these recommendations from climate science (IEA, 2019), ambition for rapid decarbonization is increasing, initially via the Paris Climate Accord (UNFCCC, 2015) and more recently via initiatives such as the European Green Deal (EU, 2019). Such a rapid decarbonization effort will require CO_2 abatement from all sectors (Bataille et al., 2018) and from existing long-lived infrastructure (Clark and Herzog, 2014). Industrial emitters are especially challenging in this respect because there are few cost-effective alternatives and decarbonization options other than post-combustion CO_2 capture generally require extensive process modifications.

Cement production is a prime example of this challenge. Currently, cement represents about 8% of global CO_2 emissions (Andrew, 2018) and, since much of the world is still industrializing, it can be safely assumed that the demand will remain robust over coming decades. There are several ways to reduce emissions from the cement industry. For example, a cement industry decarbonization roadmap from the World Business Council for Sustainable Development and the International Energy Agency expects the following contributions to emissions reduction up to 2050: 3% from increased thermal efficiency, 12% from fuel switching, 37% from reduced clinker to cement ratio, and 48% from innovative technologies (mainly CO_2 capture) (IEA, 2018).

Reduction of the clinker to cement ratio is a particularly promising early emissions reduction avenue. Supplementary cementitious materials such as fly ash and blast furnace slag are already used for this purpose, although their CO_2 mitigation potential on a full lifecycle basis is questionable (Miller, 2018). Since these materials are by-products of CO_2 intensive power plants and industries that will be gradually phased out in climate constrained pathways, the aforementioned roadmap (IEA, 2018) expects the

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contribution of these materials to cement production to decline substantially over coming decades. Instead, the reduction in clinker to cement ratio will be driven by the use of limestone filler and calcined clay. It is also possible that these clinker substitutes could play an even larger role than envisioned in this roadmap (Miller et al., 2018).

However, this 2 °C compatible roadmap only sees cement industry CO_2 emissions drop only from 2300 to 1700 ton/year. By 2050, alternatives other than CO_2 capture are expected to contribute about 200 ton/year of CO_2 emissions reductions with limited further potential (IEA, 2018). As long as clinker remains an important component of cement, deep decarbonization of the cement industry is only possible through CO_2 capture techniques, given that most of the CO_2 emissions originate from the process itself (calcium carbonate calcination required for clinker production). Although there is currently no operating CO_2 capture and storage (CCS) project in the cement industry, CCS is a proven technology with 19 large-scale facilities in operation, 4 under construction, and another 10 in advanced development, including one in the cement industry (GCCSI, 2019). A total of 260 million tons of CO_2 has been safely captured and permanently stored to date (GCCSI, 2019).

As will be reviewed in the next section, several CO_2 capture technologies have been proposed for cement production in recognition of the central role that CCS is expected to play in decarbonizing the cement industry. Following the literature review, the swing adsorption reactor cluster (SARC) CO_2 capture technology evaluated in this study will be described together with an outline of the paper.

1.1. Literature review

Cement production produces a flue gas with a relatively high CO_2 content, reducing the cost of avoiding a given fraction of produced CO_2. This, in combination with the relatively large contribution to global emissions and the fact that almost all CO_2 is delivered in a single flue gas stream, makes it one of the low-hanging fruits for implementing CCS in industry with CO_2 capture costs in the range of 26–42 $/ton (Bains et al., 2017).

For this reason, there has been considerable interest in CO_2 capture from the cement industry. A 2012 review by Kuramochi et al. (2012) compared several CO_2 capture technologies for application to cement plants using a common methodology. In the longer-term, absorption technology (37–52 €/ton) was found to be competitive with oxyfuel (43–44 €/ton) only if steam can be imported cheaply from neighbouring plants. The use of a CaO-rich feed from a power plant equipped with Calcium looping technology (Romano et al., 2013) was found to have the lowest CO_2 avoidance cost of €27/ton, but avoiding only 50% of plant emissions with the calculation assumptions by Kuramochi et al. (2012). More recently, the CEMCAP project (CEMCAP, 2018) conducted a holistic techno-economic assessment of different technologies using a common methodology. These technologies included monoethanolamine, chilled ammonia, oxyfuel, membrane assisted CO_2 liquefaction and two calcium looping configurations. Project results are summarized in separate works on technical (Voldsund et al., 2019) and economic (Gardarsdottir et al., 2019) evaluations with a more detailed account available in a deliverable report (Voldsund et al., 2018).

The comparative economic assessment yielded CO_2 avoidance costs between €42/ton for oxyfuel and €84/ton for membrane assisted CO_2 liquefaction. Oxyfuel combustion has been recognized as one of the most promising options for CO_2 capture in cement plants already in the earliest studies (IEAGHG, 2008). It was selected by the European Cement Research Academy (ECRA) for scale-up study (ECRA, 2016) and demonstration in a semi-industrial scale cement plant was recently announced by major European cement producers through an industrial joint research corporation (Beumelburg, 2019). However, the low-cost oxyfuel combustion option will require sizable modifications to several major components of the cement production plant to guarantee low ingress of ambient air in the kiln diluting the CO_2 stream (ECRA, 2016; Gerbelová et al., 2017).

One of the most competitive technologies suitable for retrofitting is the tail-end calcium looping configuration (Arias et al., 2017; Atsonios et al., 2015; Ozcan et al., 2013), which achieved a CO_2 avoidance cost of €52/ton (De Lena et al., 2019). However, retrofitting with calcium looping involves high capital cost expenditures due to the need for a heat recovery steam power plant and a large ASU (larger than the oxyfuel cement plant) and introduces additional (although manageable) complexities such as heat integration with raw meal preheaters and integration of the CaO purge from the regenerator as part of the raw meal fed to the kiln.

Monoethanolamine (IEAGHG, 2013) and chilled ammonia (Pérez-Calvo et al., 2017) processes are simpler options for retrofit, but they have higher CO_2 avoidance costs (€80/ton and €66/ton respectively (Gardarsdottir et al., 2019)) and require additional steam for solvent regeneration. Since the waste heat from the cement plant can only supply a small part of the steam by heat recovery, a natural gas boiler needs to be implemented, causing additional CO_2 emissions or requiring the capture of the CO_2 produced for steam generation. The availability of cheap steam imports from a nearby combined heat and power (CHP) plant can reduce the CO_2 avoidance cost of these technologies below €60/ton (Gardarsdottir et al., 2019). However, few cement plants are located nearby power plants that can provide such steam.

The only considered process assessed in the CEMCAP project that brings no additional complexities to the retrofitting of cement plants is membrane assisted CO_2 liquefaction (Berstad and Trædal, 2018). This technology takes the cement plant flue gas as input and captures CO_2 using only electric power from the grid. Such simplicity is highly attractive for retrofitting applications, but this technology returned uncompetitive CO_2 avoidance costs in excess of €80/ton (Gardarsdottir et al., 2019).

1.2. The swing adsorption reactor cluster

The present work focusses on a promising new technology that offers the same ease of retrofitting as membrane-assisted CO_2 liquefaction, but has the potential to achieve substantially lower CO_2 avoidance costs. This technology, called the swing adsorption reactor cluster (SARC), has been recently assessed for integration with cement plants (Cloete et al., 2019). It also requires only electricity as energy input, but achieves a lower energy penalty than all benchmark technologies aside from oxyfuel.

SARC is a low-temperature sorbent-based technology that minimizes the energy penalty through a synergistic combination of a heat pump for temperature swing and a vacuum pump for pressure swing (Fig. 1). When a vacuum is drawn, the required temperature difference between carbonation and regeneration becomes small enough that heat can be efficiently transferred from carbonation to regeneration using a heat pump (Zaabout et al., 2017). To accommodate the pressure swing, SARC uses a cluster of standalone fluidized bed reactors that dynamically cycle between carbonation and regeneration. Coordinated operation of this reactor cluster results in a steady-state processing unit.

This work will contribute to the literature by correctly positioning the novel SARC technology relative to the range of other CO_2 capture solutions available to the cement industry. The investigation is based on a bottom-up economic assessment, quantifying the

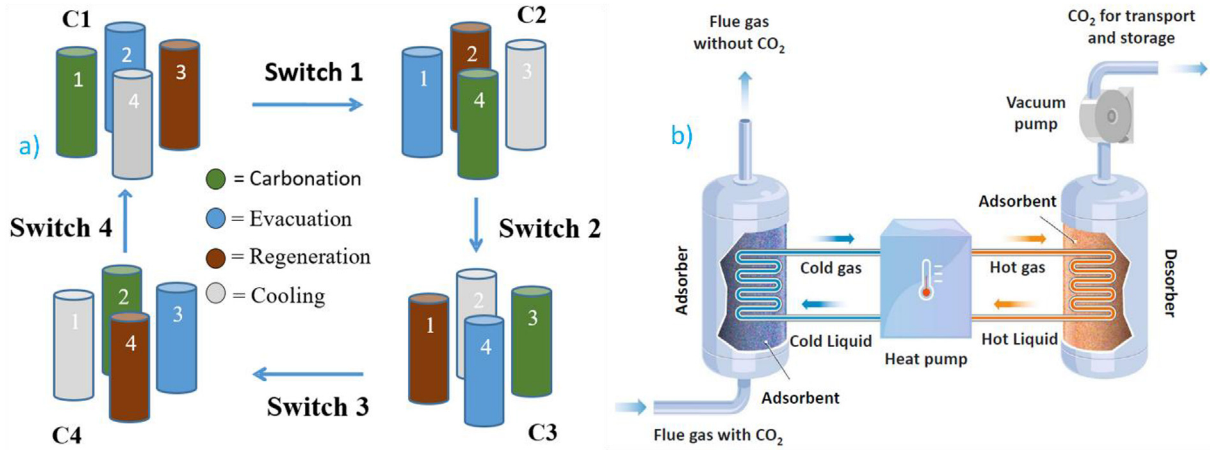

Fig. 1. SARC conceptual design: a) a cluster of SARC reactors for continuous gas stream processing; b) SARC working principle showing heat transfer from a reactor under carbonation to one under regeneration using a heat pump. Reprinted from Cloete et al. (2018), page 2, Copyright (2019), with permission from Elsevier.

CO_2 avoidance cost of SARC using a consistent methodology that allows for direct comparison to all the technologies assessed in the CEMCAP project. Key uncertainties are identified and sensitivities to important variables like the local electricity mix, capital cost escalations and fuel prices are determined. Based on these insights, the conditions under which SARC and the two most attractive benchmark technologies, oxyfuel and calcium looping, become the preferred option for CO_2 capture from cement plants are identified.

The paper is organized as follows: First, the methodology for conducting the SARC cost assessment is outlined. Subsequently, results are presented in three main parts: an optimization study to minimize SARC CO_2 avoidance costs, a sensitivity analysis to important uncertainties in the SARC cost assessment, and a comparison of SARC economic performance against several benchmark technologies over a range of relevant uncertainties. This benchmarking exercise is then used to accurately position the SARC technology relative to oxyfuel and calcium looping technologies.

2. Methodology

The methodology will be presented in two steps: a summary of the reactor and process simulations followed by a more detailed description of the method for economic assessment.

2.1. Simulations

A detailed outline of the reactor and process simulations for the SARC concept applied to cement plants is given in a recent work by the authors (Cloete et al., 2019). The same methodology is used for the simulations completed specifically for the economic assessment in the present work, so this section will only present a summary and interested readers are referred to the aforementioned work for more details. The detailed formulation of the reactor model is given in Zaabout et al. (2017).

Combined reactor and process simulations are used to simulate the SARC process performance. As detailed in the previous study (Cloete et al., 2019), a functionalized amine sorbent (Veneman et al., 2015) with known CO_2 and H_2O adsorption isotherms, density and heat capacity is used. Reactor simulations are performed using an in-house MATLAB model (Zaabout et al., 2017) that simulates the SARC reactor as a series of four CSTR's. This arrangement achieves performance between that of a CSTR and a PFR and is representative of a fluidized bed reactor with internal obstructions (including heat transfer surfaces and baffles) to limit back-mixing. Some degree of PFR behaviour is important to prevent early slippage of CO_2 and four CSTRs in series was identified as a good compromise between CO_2 capture performance and practical reactor design considerations (Zaabout et al., 2017). The four process steps of carbonation, evacuation, regeneration and cooling, illustrated in Fig. 1, are simulated sequentially in the transient reactor model and the output from each step is averaged and transferred to the process simulation.

As an example of SARC reactor behaviour, Fig. 2 shows output from the reactor model for one SARC cycle. During carbonation, CO_2 is adsorbed and heat from the exothermic carbonation reactor is continuously extracted by the heat pump to result in a gradual temperature decrease. This step is stopped when the degree of sorbent carbonation increases to the point that CO_2 slip becomes excessive. Next, a short evacuation step is conducted to vent some of the N_2-rich gases currently in the reactor to improve the purity of the CO_2 extracted in the subsequent regeneration step. Regeneration is driven by a combined pressure and temperature swing. Even though regeneration is endothermic, heat supplied by the heat pump gradually increases the bed temperature so that the sorbent releases more CO_2. In the final cooling step, the reactor temperature is reduced to a point where a sufficiently high fraction of CO_2 can be adsorbed at the start of the carbonation step initiating the next SARC cycle.

For each case, the model adjusts the heat pump working fluid condensation temperature to achieve 90% CO_2 capture and the evacuation pressure to achieve 96% CO_2 purity. A higher condensation temperature will increase the maximum temperature in the regeneration step (at the expense of greater heat pump power consumption), releasing more CO_2 so that a greater fraction of CO_2 can be captured in the next carbonation step. A lower evacuation pressure withdraws more N_2-rich gases from the reactor in the evacuation step so that the CO_2 purity from the subsequent regeneration step increases (at the expense of a small amount of CO_2 being vented with the N_2-rich gases). The resulting working fluid condensation temperature achieving 90% CO_2 capture and evacuation pump pressure achieving 96% CO_2 purity are passed to the process model together with the total amount of heat transferred from carbonation to regeneration by the heat pump. Using this information, in combination with the fixed working fluid evaporation temperature of 55 °C, the process model can then calculate the heat pump and evacuation pump power consumption for use in the economic assessment.

In this work, new simulations were carried out at different levels

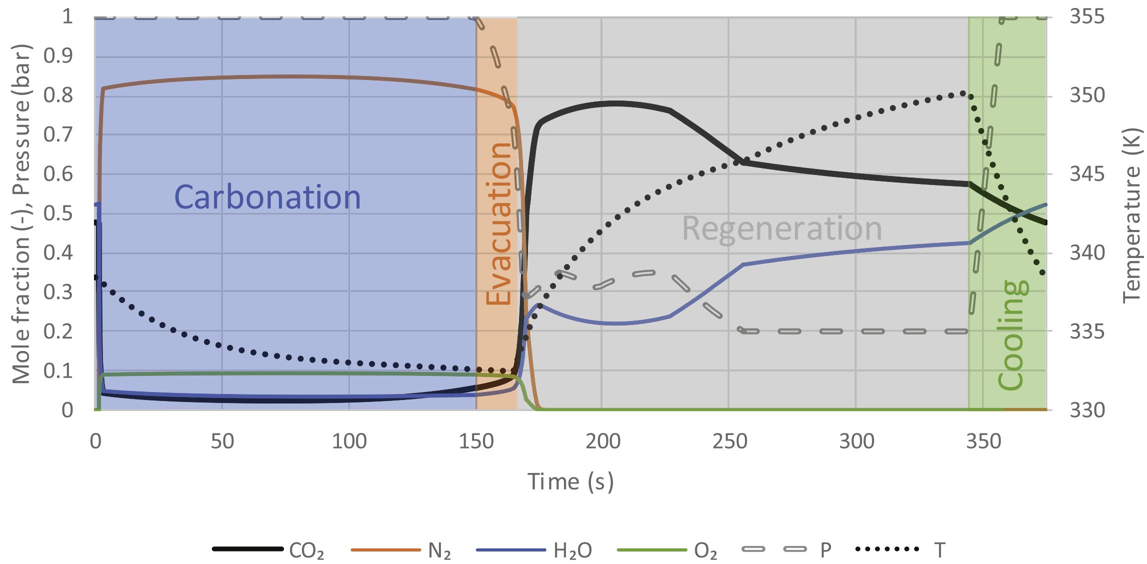

Fig. 2. Behaviour of the SARC reactor over a single cycle in the base case considered in this study.

of heat transfer surface area and regeneration pressure to investigate the trade-off between capital costs and process efficiency. Specifically, a larger heat transfer surface area will allow for a smaller temperature difference between the sorbent and the working fluid, thereby reducing the difference between the heat pump working fluid evaporation and condensation temperatures required to achieve a given temperature swing in the reactor. This increases heat pump efficiency at the expense of higher heat exchanger capital costs. A higher regeneration pressure means a smaller pressure swing is imposed, saving vacuum pump power consumption and capital costs. However, this requires a larger temperature swing to achieve the same degree of CO_2 capture, which increases the heat pump compressor power consumption and capital costs. These are the most important economic optimization parameters involved in the SARC process.

Each simulation was completed for a cluster of 25 reactors. At any given point in time, 10 of these reactors are operating in carbonation, 1 in evacuation, 12 in regeneration and 2 in cooling. For each individual reactor, this results in the distribution of step times indicated in Fig. 2. Reactors were sized assuming a fluidization velocity of 1 m/s in the carbonation step (just below the onset of turbulent fluidization when typical 150 μm particles are used (Bi and Grace, 1995), which meant that each reactor was 3.46 m in diameter. The reactor height varied between the cases depending on the heat transfer surface area to keep the amount of sorbent in the reactor constant at 15 tons. Cases with more surface area included more tube volume in the reactor, requiring a taller reactor to accommodate the same sorbent loading. In each case, freeboard volume equivalent to 80% of the height of the static bed and tube volume was added to the height of the reactor to prevent excessive particle elutriation (Zaabout et al., 2017). An increase in heat transfer surface area also increased the heat capacity of the reactor, requiring more heat transfer to achieve a given temperature swing. More details on reactor height and effective heat capacities for the different cases are given in Section 3.1.

Using the input data from the reactor model, process modelling is completed using the in-house code GS (Gecos, 2014) and ASPEN Plus (AspenTech, 2016). As shown in Fig. 3, the only point of connection between the cement plant and the SARC process is the flue gas stream that is boosted by the flue gas fan to the carbonation step of the SARC process. The CO_2-lean flue gas emerging from the carbonation step is then sent to the stack together with the small amount of N_2-rich gas extracted during the short evacuation step. The heat pump extracts heat from the exothermic carbonation reaction and transfers it to the endothermic regeneration step where a vacuum is drawn and the bed temperature is increased to release CO_2 from the sorbent. The CO_2 is extracted by the vacuum pump in three intercooled stages and subsequently compressed in three intercooled stages. A small fraction of the extracted CO_2 stream is recycled using the CO_2 recycle fan to ensure good fluidization in the regeneration step.

The purpose of the process modelling is to determine the power consumption of the CO_2 compressors, vacuum pumps (large pump for regeneration and small pump for evacuation), heat pump compressor, flue gas fan, CO_2 recycle fan, and additional auxiliary consumption. This information is then used in the economic assessment to size the different equipment and calculate the cost of the electricity required to run the SARC process.

2.2. Economic assessment

To ensure accurate benchmarking against alternative options for CO_2 capture from cement plants, this study employs the same economic assessment methodology as recently published from the Horizon 2020 project, CEMCAP (Gardarsdottir et al., 2019). The key performance indicators (KPIs) are calculated identically to the CEMCAP project using the Excel model available online (De Lena et al., 2018). The KPIs are summarized below.

First, the cost of clinker (COC) is calculated through the annualized contributions of capital expenditures, fuel, raw materials, electricity and other operating and maintenance costs (Equation (1)). Capital costs are annualized using the assumptions detailed in Table 1. A detailed description of each individual cost component is provided later in this section.

$$COC = C_{cap} + C_{fuel} + C_{RM} + C_{el} + C_{O\&M} \quad \text{Equation 1}$$

Next, the equivalent CO_2 emissions are calculated, summing direct and indirect emissions (Equation (2)). In the case of SARC, indirect emissions originate from imported electricity (the carbon intensity of imported electricity is assumed to be 262 kg/MWh as a reasonable European average (Voldsund et al., 2018)). For consistency, CO_2 emissions are expressed in tons of CO_2 per ton of clinker.

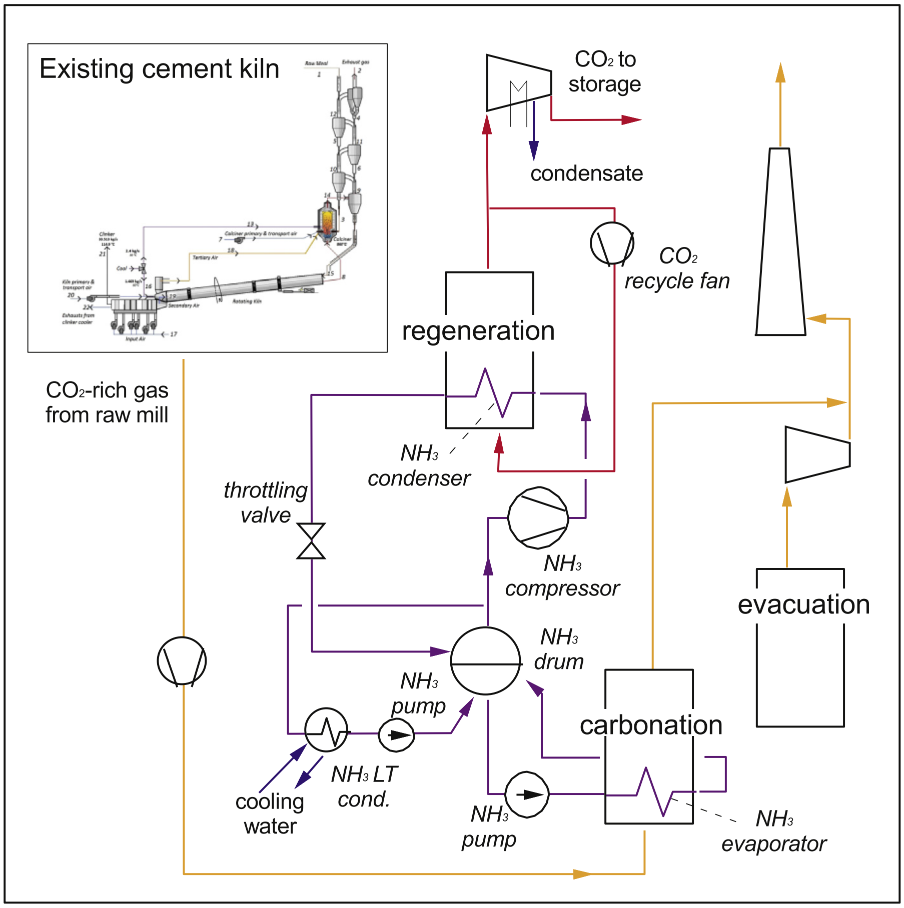

Fig. 3. Schematic of the SARC concept for CO_2 capture from a cement plant originally proposed in (Cloete et al., 2019). The reactor under cooling step is not shown since it behaves the same as the carbonation step.

Table 1
Key assumptions in calculating the annualized capital cost.

Capacity factor (%)	91.3
Economic lifetime (years)	25
Cement plant construction time (years)	2
Construction expenditure breakdown (%)	50/50
SARC plant construction time (years)	3
Construction expenditure breakdown (%)	40/30/30
Discount rate (%)	8

$$E_{eq} = E_{dir} + E_{el} \quad \text{Equation 2}$$

Using these two KPIs, the most important comparison metric, the CO_2 avoidance cost (*CAC*) can be calculated as in Equation (3), where "*ref*" indicates the reference cement plant without CO_2 capture. Details about the reference plant can be found in Gardarsdottir et al. (2019).

$$CAC = \frac{COC - COC_{ref}}{E_{eq.ref} - E_{eq}} \quad \text{Equation 3}$$

2.2.1. Capital costs

The SARC process evaluated in this study consists of a cluster of 25 fluidized bed reactors. Each reactor consists of a simple cylindrical process vessel that must be reinforced to withstand vacuum operation. Heat transfer tubing is installed inside each vessel to carry the heat pump working fluid.

In addition, the SARC process requires two vacuum pumps: one main pump that operates in the regeneration step and one small pump that operates in the evacuation step. The CO_2 stream extracted by the main vacuum pump must subsequently be compressed for transport and storage, requiring a CO_2 compressor train.

Furthermore, the heat pump requires an ammonia compressor and storage vessel. Two blowers are also required to overcome the pressure drop in the reactor: one large blower for driving the flue gas through the SARC carbonation and cooling steps and another small blower for a small recycle stream fed to the regeneration step.

In accordance with the CEMCAP methodology, substantial process contingencies are added to the capital cost estimates of different components as outlined in Table 2. The large contingency of the SARC reactors signifies its low level of development. Other components are at a higher level of technological readiness and are therefore assigned a smaller contingency. Furthermore, an additional process contingency is added to all components to account for any missing detail in the equipment list.

Capital costs of different components are primarily estimated using cost correlations from Turton et al. (2008). This methodology provides cost correlations for purchased costs of equipment constructed from carbon steel for atmospheric pressure operation. Subsequently, additional correlations are used to add installation costs and adjust for cost increases from pressurized (or vacuum) operation and the use of more expensive materials. The selected materials and pressures for the different process components are

Table 2
Process contingencies employed in the SARC capital cost estimation.

Reactor vessels and heat exchange surfaces	40%
Vacuum pumps, compressors, blowers and NH_3 storage vessel	20%
Additional contingency from low detail in the equipment list (all components)	12%

summarized in Table 3 and the different cost correlations used are given in the appendix.

Costs of two process components were estimated from other sources. First, the cost of the CO_2 compression train was taken directly from the CEMCAP project. Second, the installed cost of a vacuum pump was estimated from the generalized vacuum pump correlation provided by Woods (2007). This cost was then adjusted from carbon steel to stainless steel using material factors from Turton et al. (2008). All costs were adjusted to 2014 Euros using the CEPCI index and an exchange rate of 1.2 $/€.

The cost of the reactor vessels was increased by 50% to account for additional elements such as the gas distributor, valves and gas-solid separators. Furthermore, the cost of the heat exchange surfaces in each SARC reactor was taken as half of the cost of a conventional shell and tube heat exchanger because only the heat exchanger tubing is required for the SARC application. This assumption was tested by using the heat exchanger cost correlation (Turton et al., 2008) to estimate the installed cost of 1) a heat exchanger with carbon steel tubes and shell, 2) a heat exchanger with stainless steel tubes and a carbon steel shell, and 3) a heat exchanger with stainless steel tubes and shell. The cost increase from case 1 to case 2 was almost identical to the increase from case 2 to case 3, indicating that the costs associated with the shell and tubes are similar in magnitude.

SARC reactors require large heat exchange surface areas and large vacuum pump flowrates, extending beyond the range of validity of available cost correlations. In the base case, it was assumed that no further economies of scale are possible beyond the upper limit of the given range of validity: 1000 m^2 for the heat exchanger (Turton et al., 2008) and 500 (kg/h)/kPa (mass flowrate in kg/h divided by the absolute pressure of the vacuum in kPa) for the vacuum pump (Woods, 2007). The implications of continued economies of scale is explored in a later section.

An important simplifying assumption was required to calculate the vacuum pump costs. Since the correlation is valid for a single stage pump, it is not ideally suited to the present case with a train of three vacuum pumps in series. On the one hand, simply assuming a single vacuum pump in the cost assessments would result in an under-estimation of the cost because one large pump will be cheaper than three smaller pumps in series. On the other hand, applying the correlation to all three pumps in series would over-estimate the costs because the individual pumps in series operating over a smaller pressure ratio would be cheaper than the single stage pumps assumed in the correlation.

A convenient middle ground between these two bounds was found by simply assuming three single stage pumps operating in parallel. It was checked that this simplifying assumption generally resulted in costs roughly half-way between the two bounds outlined above. For this reason, cost calculations for the cases without continued economies of scale assumed multiple parallel pumps with a capacity of 500 (kg/h)/kPa, and the cases with continued economies of scale assumed three equally sized parallel pumps. In general, the vacuum pump cost is an important uncertainty in this assessment and a dedicated section is presented in the results and discussion to further explore this topic.

Table 3
Material and pressure assumptions for different process units.

Equipment	Material	Pressure (bar)
Reactor vessel	Carbon steel	0.1
Heat exchanger tubes	Carbon steel	30
NH_3 buffer vessel	Carbon steel	30
Vacuum pumps	Stainless steel	–
CO_2 compressors	Stainless steel	–
Other compressors and blowers	Carbon steel	–

Finally, the initial loading of sorbent in the SARC reactors was also included in the capital cost. A cost of €15/kg was assumed for large scale production of the PEI sorbent used in the reactor simulations based on discussions with potential future suppliers.

Following these calculations, a typical breakdown of capital costs is presented in Fig. 4.

Adding up the installed equipment costs yields the total direct cost (TDC). The TDC is then increased by adding indirect costs (14% of TDC), owner's costs (7% of TDC) and a project contingency (15% of TDC), yielding the total plant cost (TPC).

2.2.2. Operating costs

Out of the remaining components in Equation (1), only the electricity and operating and maintenance costs are relevant to SARC. Electricity consumption is determined in the process simulations described earlier. From this result, the electricity cost is calculated by assuming an electricity price of €58.1/MWh (Gardarsdottir et al., 2019). This price is representative for large industrial electricity consumers in Europe (CREG, 2019).

Operating and maintenance costs consist of routine maintenance assumed as 2.5% of TPC per year and insurance of 2% of TPC per year (Gardarsdottir et al., 2019). In addition, sorbent replacement costs are added by assuming a 2-year operating lifetime and a cost of €15/kg.

SARC has no effect on the operating costs of the cement plant, which are presented in Gardarsdottir et al. (2019).

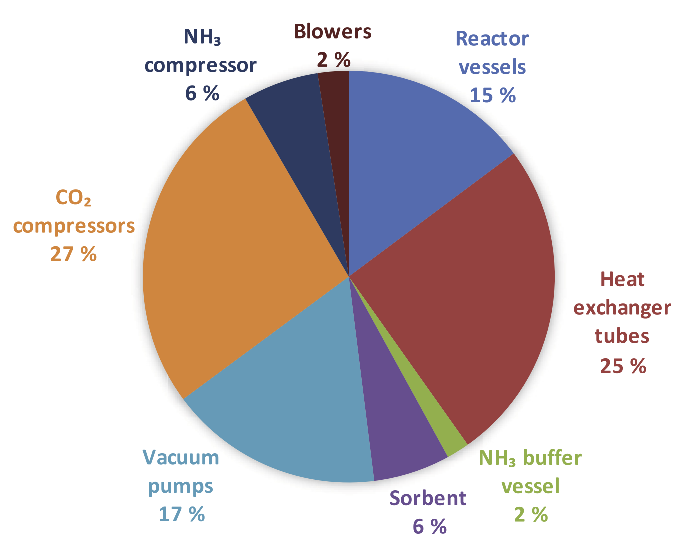

Fig. 4. Capital cost breakdown (including process contingencies) for the central case considered in this study.

3. Results and discussion

Results will be presented in six sections. First, the sensitivity of SARC economic performance to design choices regarding the heat exchange surface area and vacuum pressure will be discussed. The next three sections will discuss more optimistic scenarios: 1) specialized heat exchangers and vacuum pumps for the SARC process allow for continued economies of scale, 2) cheaper large-scale vacuum pumps allow for optimal performance at lower regeneration pressures, and 3) incentives for electrification using renewables substantially reduce effective electricity costs and carbon intensity. Finally, a detailed benchmarking and sensitivity study will be presented in two parts: 1) comparison of the economic performance of SARC to benchmark technologies and 2) positioning of SARC relative to the two most attractive benchmark technologies: oxyfuel and tail-end calcium looping.

3.1. Economic performance of SARC

The SARC process offers some interesting trade-offs. First, there is a trade-off between vacuum swing and temperature swing. Stronger vacuums during regeneration allow for a smaller temperature swing, which increases heat pump efficiency. This results in a trade-off between vacuum pump and heat pump power consumption.

Another important trade-off is related to the available heat exchange surface area within the reactor. Larger surface areas increase the plant capital cost, but allow for a smaller temperature difference between the reactor and the heat pump working fluid, thus lowering the required difference between condensation and evaporation temperatures to increase heat pump efficiency.

To explore these trade-offs, simulations were completed in a 2-dimensional parameter space spanned by heat exchanger surface area and vacuum pressure using the nine cases summarized in Table 4. The levels of the nine simulations were chosen according to a central composite design methodology to ensure optimal spacing of the cases throughout the parameter space (Montgomery, 2001).

As introduced in Section 2.1, the amount of heat exchange surface area included had two other important effects. First, inclusion of more tubing increased the thermal mass inside each reactor, requiring more heat transfer to achieve a given temperature swing. This effect was accounted for by changing the effective specific heat capacity (c_p) of the sorbent in each case as shown in Table 4, assuming a tube wall thickness of 0.9 mm and a reactor wall thickness of 20 mm. For example, the central case (Case 9) almost doubled the c_p of the sorbent (1500 J/kg.K) to 2951 J/kg.K by accounting for the heat capacity of the 44 tons of steel ($c_p = 500$ J/kg.K) used in each reactor next to the 15 tons of sorbent. Second, more tubes took up additional space inside each reactor. To standardize the simulations, the amount of active reactor volume was kept constant, requiring the reactor height to be changed with a change in the volume fraction occupied by the tubing (assumed to be 6 mm outer diameter). This influenced the pressure drop and the power consumption of the flue gas blower.

The process simulation results and the installed costs of the main plant components are shown in Fig. 5 for each case, and the Excel model used to generate these results is attached as supplementary material. Substantial variation can be observed between cases. In Fig. 5a, the cases with higher regeneration pressures (e.g. Case 6) have lower vacuum pump consumption and greater heat pump consumption. Heat pump consumption is also influenced by the amount of heat transfer surface area included in the reactors. As expected, Fig. 5b shows higher costs for cases with large heat exchange surface areas (e.g. Case 8) and strong vacuums (e.g. Case 5).

It is also clear from Fig. 5 that cases with higher capital costs (e.g. Case 5) generally have lower power consumption. Fig. 6 better illustrates this trade-off, showing that low regeneration pressures and large heat exchange surface areas generally result in low electricity consumption, but large capital costs. This trade-off leads to relatively small variations in COC and CAC across the parameter space. As expected, the trends in COC and CAC are similar, although CAC is somewhat more sensitive to lower heat exchange areas because these cases consume more electricity, which leads to more indirect emissions and thus lower CO_2 avoidance.

Fig. 6 also shows that optimal performance is achieved at moderate heat exchange surface areas and low regeneration pressures, although excessively low regeneration pressures (Case 5 in Table 4) also leads to worse economic performance. Fig. 5 shows that Case 5 is the most efficient, but also has the highest capital costs due to the large costs associated with the vacuum pump. Overall, Case 1 achieved the lowest overall COC and CAC at 99.5 and 52.4 €/ton respectively.

CAC is broken down into its major components for each case in Fig. 7. Costs are relatively evenly distributed between fixed (capital and fixed O&M) and variable (electricity and sorbent) costs. It is also clear that lower capital costs generally result in higher electricity consumption and vice versa.

3.2. Continued economies of scale

As mentioned in Section 2.2.1, the economic assessment in the previous section considered no economies of scale beyond 1000 m^2 of heat exchange area and 500 (kg/h)/kPa of vacuum pump flow. However, if SARC is deployed on an industrial scale, it is likely that specialized equipment will become available capitalizing on continued economies of scale beyond these limits. This section will therefore illustrate the effect of continued economies of scale on CAC.

For the heat exchanger, it is assumed that economies of scale can extend up to the maximum size considered in this study: 6828 m^2. As discussed in Section 2.2.1, the vacuum pump cost assessment was done for three parallel single stage pumps, so it was assumed that economies of scale extend up to a maximum flowrate of 6338 (kg/h)/kPa.

A comparison of Fig. 8a to Fig. 5b shows that continued economies of scale substantially reduced the capital costs, particularly those of the vacuum pumps, which uses a scaling exponent of 0.64 (Woods, 2007). As a result, the cases with a low regeneration pressure become substantially more attractive. Fig. 8b shows that CAC continues to decrease with regeneration pressure if the large

Table 4
Summary of the nine cases considered for this study.

Case	1	2	3	4	5	6	7	8	9
Regeneration pressure (bar)	0.100	0.300	0.100	0.300	0.059	0.341	0.200	0.200	0.200
Heat exchange area (m^2)	2000	2000	6000	6000	4000	4000	1172	6828	4000
Reactor height (m)	6.010	6.010	7.158	7.158	6.584	6.584	5.772	7.396	6.584
Tube volume fraction	0.096	0.096	0.241	0.241	0.174	0.174	0.058	0.265	0.174
Effective c_p (J/kg.K)	2435	2435	3468	3468	2951	2951	2221	3681	2951

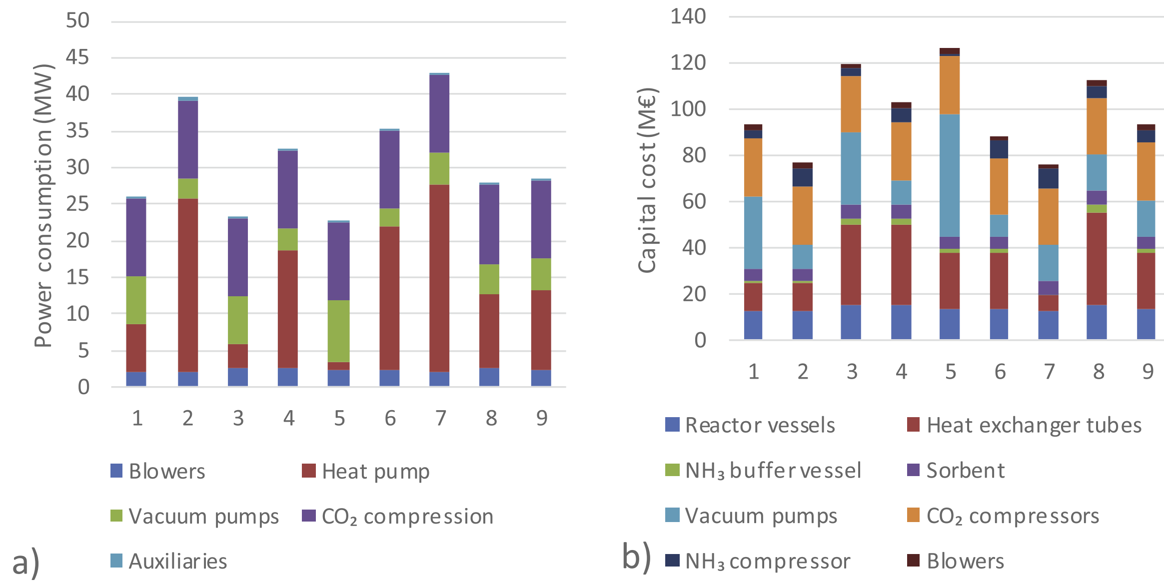

Fig. 5. A breakdown of power consumption (a) and capital costs (b) for the nine cases listed in Table 4.

vacuum pumps required at the lower end of the range can benefit from continued economies of scale. This is explored further in the following section.

Case 1 remained the most economic case, returning a COC and CAC of 95.7 and 47.0 €/ton respectively (11% reduction in CAC). For perspective, the vacuum pumps in this case need to be 7x larger than the upper limit of the cost correlation employed, whereas the heat exchangers are only 2x larger.

3.3. Lower regeneration pressures

As shown in Fig. 8b, CAC continues to decrease with regeneration pressure when continued economies of scale are assumed. To date, there are limited applications that require very large vacuum pumps and, in anticipation of a sizable market for large machines created by vacuum swing adsorption CO_2 capture applications, it is worthwhile exploring this dynamic further.

The vacuum pump cost correlation deployed in the results outlined above is scaled based on scaling laws derived from existing machines, which are mostly positive displacement machines suitable for much smaller size compared to those needed for an industrial CO_2 capture plant. When considering large vacuum pumps for the SARC system, multi-stage turbo-compressors are likely to be used. Therefore, vacuum pump costs have also been calculated according to compressor cost correlations, which scale according to the compressor power consumption.

Thus, this section will investigate the effect of these different options for vacuum pump cost estimations at lower regenerations pressures comprising of four cases: 0.1, 0.075, 0.05 and 0.025 bar. Aside from the vacuum pressure, all other settings are the same as for Case 1 in Table 4, given that this case generally gave the best results. The four vacuum pump cost correlations are as follows:

1. Flow - no EoS. In this case, the flowrate cost correlation (Woods, 2007) is used without any economies of scale beyond 500 (kg/h)/kPa.
2. Flow - EoS. The same cost correlation is used, but with continued economies of scale, assuming three equally sized parallel vacuum pumps.
3. Power - no EoS. In this case, the standard compressor cost correlation (Turton et al., 2008) is used with no economies of scale beyond a flowrate of 500 (kg/h)/kPa.
4. Power - EoS. The same standard compressor cost correlation with continued economies of scale is applied to three vacuum pumps with equal power consumption.

It should also be mentioned that, in the cases with 0.025 bar regeneration pressure, the heat pump is no longer required since the regeneration temperature falls below the carbonation temperature. This allows the reaction heat to be moved by a water pump. In this case, the heat pump power consumption and ammonia pump compressor cost are set to zero.

The results of this study are summarized in Fig. 9. A minimum power consumption is achieved at a regeneration pressure of 0.05 bar (Fig. 9a), while capital costs continue to increase with decreasing regeneration pressure (Fig. 9b), mainly due to rising vacuum pump costs.

The effect of the four different vacuum pump cost correlations is shown in Fig. 10. Clearly, the results returned by the different correlations vary over a wide range, particularly at low regeneration pressures. Fig. 10a shows that the highest vacuum pump cost can be as much as 6x higher than the lowest vacuum pump cost at a regeneration pressure of 0.025 bar. Continued economies of scale have a large positive effect.

Since the vacuum pump is just one element of the total system cost, the effect on CAC shown in Fig. 10b is not as dramatic. A large effect is only observed in the case of the flowrate cost correlation without economies of scale (Flow − No EoS). This case shows substantial increases in CAC with lower regeneration pressures as could be anticipated from Fig. 6.

For the remaining three vacuum pump cost correlations, the effect is smaller, with CAC varying between 45.4 and 53.2 €/ton over the 12 cases. The optimal regeneration pressure in these three cost correlations is mildly dependent on the regeneration pressure in the investigated range and is either 0.05 or 0.075 bar, with optimal CAC varying between 45.4 and 49.2 €/ton.

These results suggest that more detailed engineering studies regarding large-scale vacuum pump design and cost are required. If continued economies of scale are possible for large vacuum pumps,

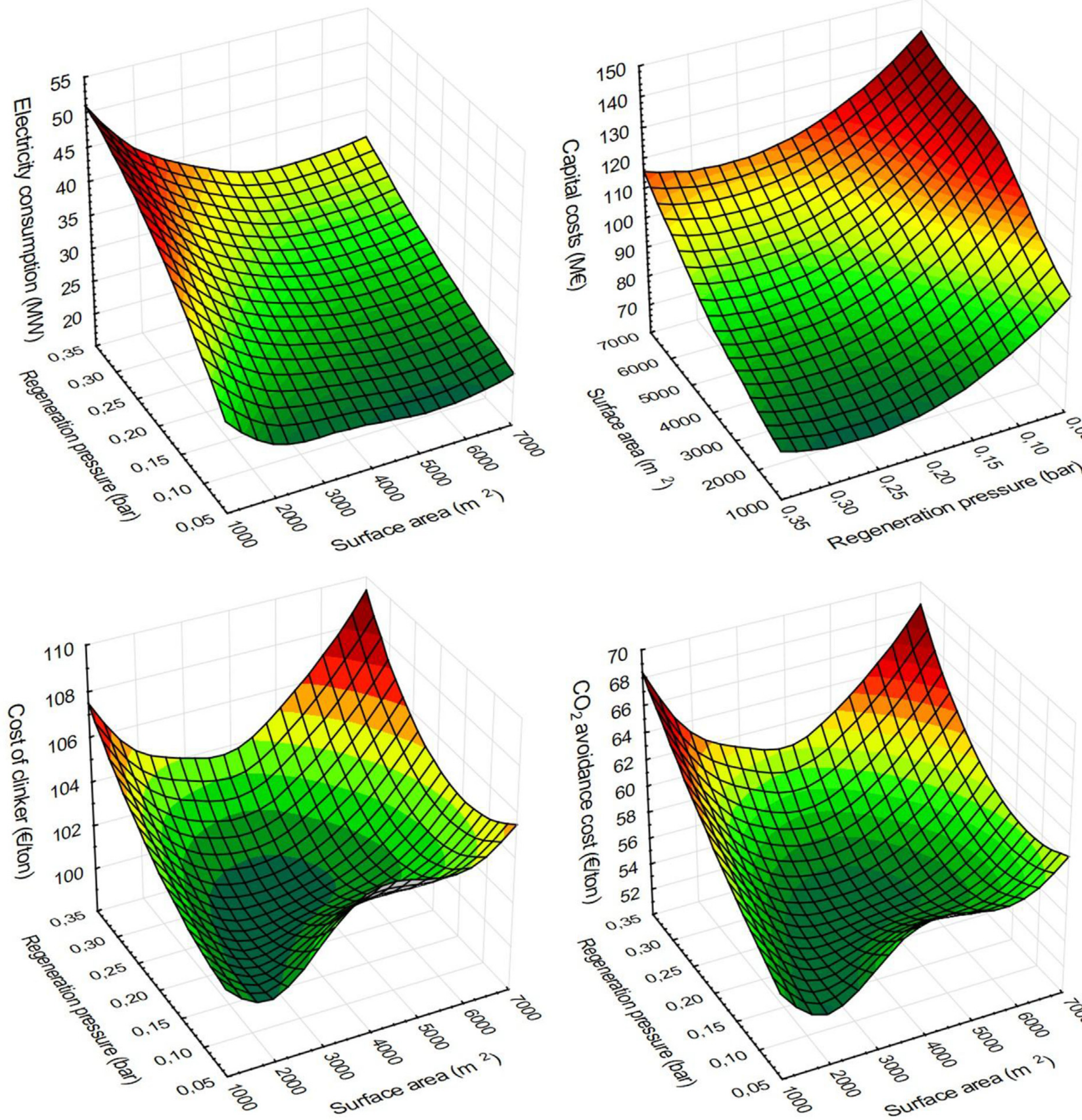

Fig. 6. Spline surface plots of the electricity consumption, plant capital costs, cost of clinker (COC) and CO_2 avoidance cost (CAC).

significant improvements in CAC are possible. Similarly, if vacuum pump costs approach compressor costs (scaled according to power consumption), the economic attractiveness of SARC increases further.

3.4. Electrification incentives

Electrification of industry through renewable energy is a topic of considerable interest at present and may attract significant incentives over coming decades. As a technology that consumes only electricity, SARC will benefit strongly from lower electricity prices and CO_2 intensities. Lower electricity prices will directly reduce the operating costs of the plant, whereas lower CO_2 intensities will increase overall CO_2 avoidance, thus lowering CAC (Equation (3)).

To illustrate this effect, this section will repeat the study (without continued economies of scale) in two steps: 1) lowering the electricity price from €58.1/kWh to €40/kWh and 2) lowering the CO_2 intensity from 262 kg/MWh to 0 kg/MWh. Incidentally, this is indicative of the situation in Norway where cheap hydropower supplies almost all electricity.

Fig. 11a shows a substantial change in the shape of the response of CAC relative to Fig. 6. The change is particularly large at low heat exchanger surface areas where the electricity consumption by the heat pump is larger due to the larger difference between working fluid condensation and evaporation temperatures. With lower electricity prices and zero indirect emissions (Fig. 11b), this trade of higher electricity consumption for lower capital costs by reducing the heat exchanger surface area becomes more attractive.

Case 1 remains the most attractive case with CAC values of €46.8/ton for the case with a lower electricity price and €43.3/ton for the case with an additional reduction in CO_2 intensity.

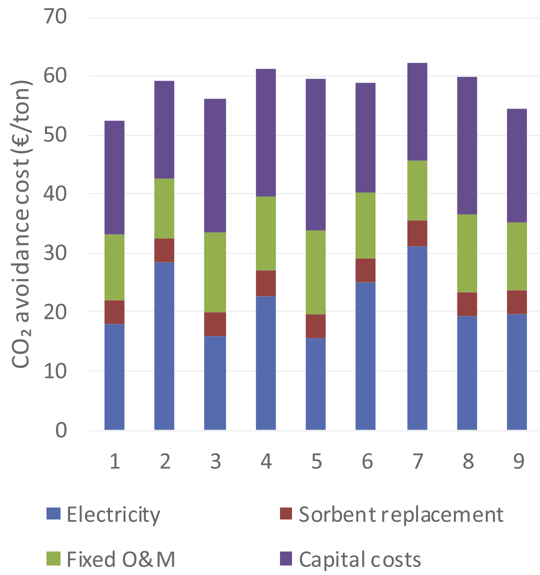

Fig. 7. Breakdown of CO_2 avoidance cost for the nine cases listed in Table 4.

3.5. Benchmarking and sensitivity analysis

The SARC concept was evaluated using the same methodology as used in the CEMCAP project to allow for easy benchmarking to competing technologies. Fig. 12 shows that SARC (with the configuration of Case 1 in Table 4 and without continued economies of scale) compares well against alternatives with only oxyfuel technology achieving a consistently lower CAC. SARC returns almost identical costs to the CaL tail-end configuration in the base case and outperforms the other technologies.

Since SARC consumes energy only in the form of electricity, it shows no sensitivity to coal or steam prices. CaL is the only technology consuming coal, whereas CAP and MEA technologies consume steam. Thus, CaL outperforms SARC at low coal prices, but the situation is reversed at high coal prices. The competitiveness of CAP and MEA improves with low steam prices (which will require integration with another process producing excess steam), but their costs remain above that of SARC.

The sensitivity to electricity prices is the most interesting, particularly in the comparison between SARC and CaL. Whereas SARC consumes electricity to capture CO_2, CaL produces electricity in a steam cycle using the heat released in the regenerator. This electricity production can cover part of the cement plant's demand. For this reason, SARC benefits from low electricity prices, while CaL tail-end benefits from high electricity prices. The net effect is substantial, with a swing of almost €20/ton in CAC between SARC and CaL tail-end over the range investigated.

Since SARC imposes a larger energy penalty than oxyfuel technology, the difference between the CAC of these two technologies also increases moderately with electricity price. The other technologies also consume some electricity and therefore remain more expensive than SARC over the range investigated. Even so, these results suggest that SARC is unlikely to be the preferred option in regions where electricity prices paid by large industrial consumers exceed the upper bound in this sensitivity analysis (€87.2/MWh).

Lastly, the capital cost sensitivity of SARC is lower than most other technologies, with only oxyfuel and MEA technologies displaying a similar cost sensitivity. Although the SARC reactors are large, they are relatively simple and can be constructed from carbon steel. The more complex high temperature reactor configuration of CaL tail-end combined with the addition of a steam power cycle requires a larger capital investment and therefore shows greater sensitivity to capital cost variations.

Fig. 13 shows that SARC has a relatively high sensitivity to CO_2 price compared to other CO_2 capture technologies. This is due to 10% direct emissions (90% CO_2 capture) as well as significant indirect emissions from electricity consumption. With an electricity CO_2 intensity of 262 kg/MWh, the total CO_2 emissions are divided about 60/40 between direct and indirect emissions.

In a future scenario with very high CO_2 prices, there will be great pressure on electricity producers to reduce CO_2 emissions, implying that the indirect emission challenge faced by SARC will diminish. SARC can also be configured to capture higher fractions of the CO_2 in the flue gas, although this will require a larger temperature and/

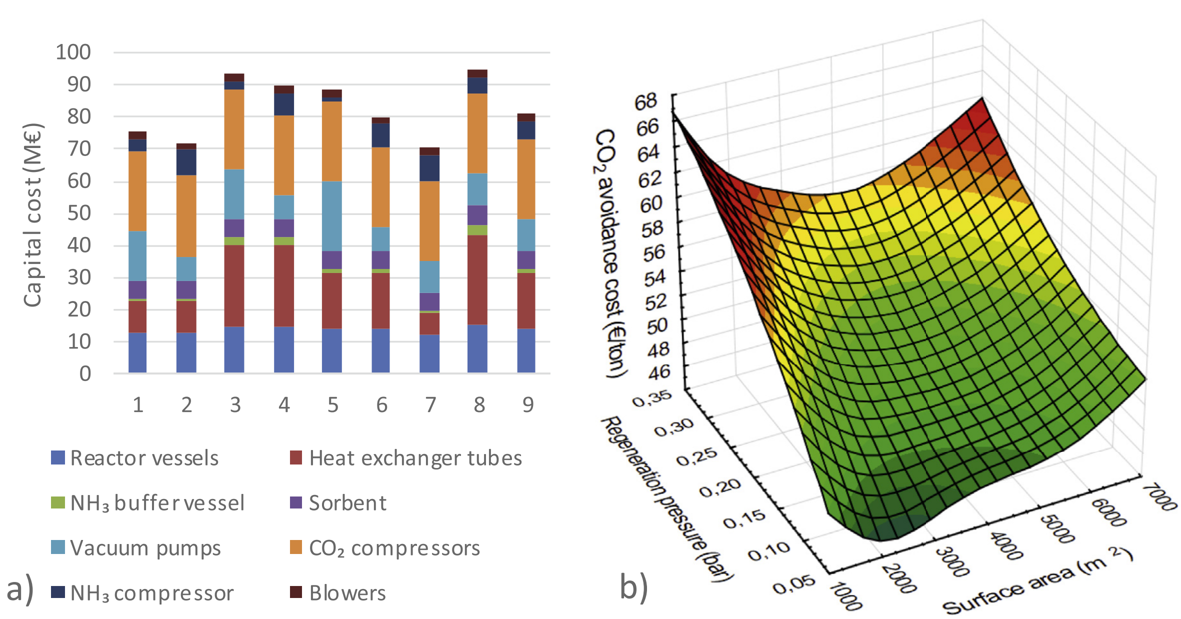

Fig. 8. Capital cost breakdown (a) and CO_2 avoidance cost (b) variation over the nine cases listed in Table 4.

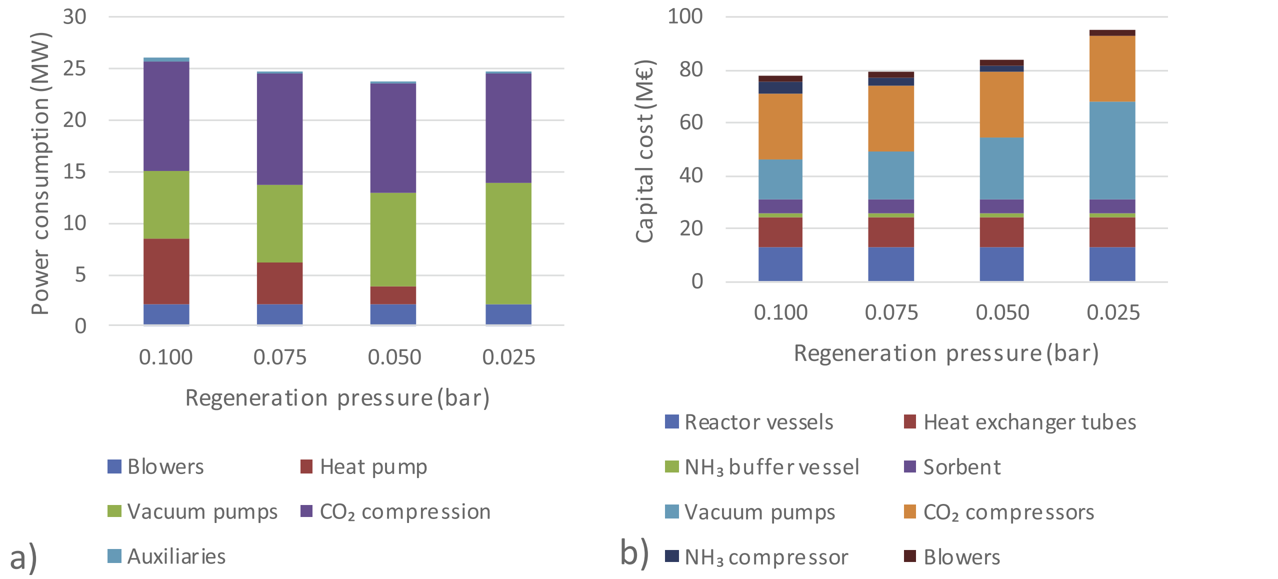

Fig. 9. A breakdown of power consumption (a) and capital costs (b) of the four different regeneration pressures. Capital costs are reported for the flowrate vacuum pump cost correlation with continued economies of scale (Flow − EoS).

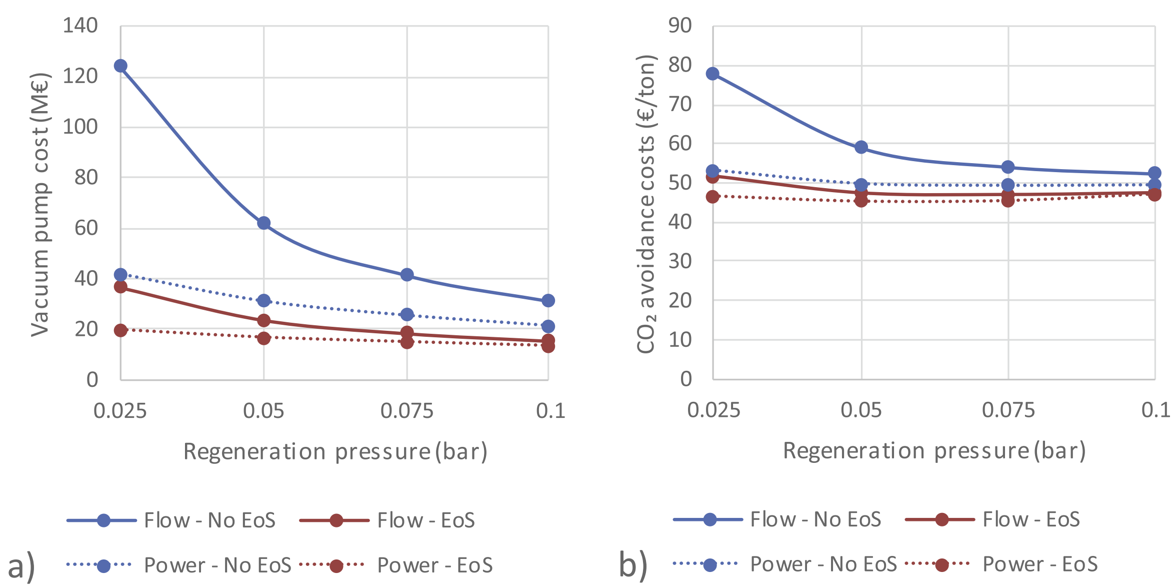

Fig. 10. The effect of vacuum pump cost correlation on vacuum pump capital cost (a) and CAC (b).

or pressure swing, thus increasing electricity consumption.

3.6. Positioning of SARC technology

From this analysis, the three most promising technologies for CO_2 avoidance from cement production are oxyfuel, SARC and tail-end calcium looping. The key parameters influencing the relative attractiveness of these three technologies are the local electricity mix and the importance of retrofittability. As illustrated in Fig. 14, oxyfuel technology is the preferred option when retrofittability is not of high importance, e.g. for construction of a new cement plant or for existing plants scheduled for a long shutdown for modernization or installation of other new equipment. Tail-end calcium looping is preferred when local electricity is expensive and/or carbon-intensive, whereas SARC and oxyfuel technology benefit from cheap and/or low-carbon electricity. SARC is preferred when retrofittability is of high importance, given that it can be constructed while the host process is operating as usual and subsequently connected with almost no downtime in cement production.

Qualitatively, the angles of the three numbered separating lines in Fig. 14 can be explained as follows: Line 1 indicates that calcium looping becomes competitive against oxyfuel at lower electricity prices as retrofittability becomes more important, given that it has a lower impact on the host process. Line 2 shows that oxyfuel becomes more competitive against SARC as the electricity price increases because it consumes less electricity per unit of CO_2 captured. Line 3 shows that SARC becomes competitive against calcium looping at higher electricity prices as the importance of retrofittability increases because it has no impact on the host process.

The quantitative positioning of these lines depends on several

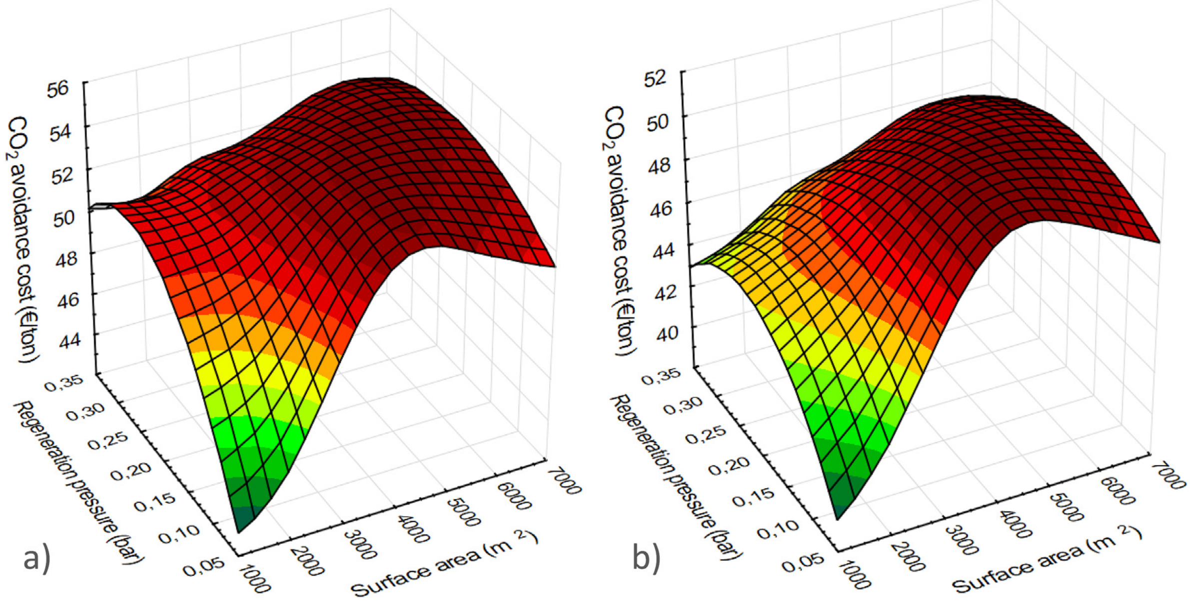

Fig. 11. CO_2 avoidance costs for a reduced electricity price of €40/MWh (a) and an additional reduction in CO_2 intensity to 0 kg/MWh (b).

factors. From the comparative performance presented in Fig. 12, it can be stated that the point where lines 1, 2 and 3 in Fig. 14 meet is roughly equivalent to the electricity mix assumed in this study (a cost of €58.1/MWh and a CO_2 emissions intensity of 262 kg/MWh) and levelized cost of retrofitting with oxyfuel technology (e.g. the cost of shutting down production for an extended time period) roughly equal to €10/ton of CO_2. However, future quantitative comparative studies should be done in greater detail on a case-by-case basis considering the six key uncertainties listed below.

The first important uncertainty is the technical development of these three CO_2 capture technologies. All three are at relatively low technology readiness levels at present, posing a risk of unexpected challenges during scale-up. SARC is well positioned in this respect due to the simplicity of its standalone bubbling fluidized bed reactors and its low operating temperature, although continuous operation of the reactor cluster and sufficient chemical and mechanical resistance of the sorbent material need to be demonstrated. Tail-end calcium looping is at relatively higher TRL, as it has already been demonstrated at industrially relevant conditions and operates under conditions (e.g. solids circulation rate, reactor temperature) similar to commercial circulating fluidized bed boilers. Oxyfuel requires demonstration at semi-industrial scale to validate the concept and demonstrate that low air infiltration can be maintained for sufficiently long continuous operating time.

Second, uncertainties in the cost assessments of these technologies can significantly influence their relative competitiveness. For example, Sections 3.2 and 3.3 showed that greater economies of scale in key equipment, particularly the vacuum pump, can allow SARC to achieve CO_2 avoidance costs approaching that of oxyfuel combustion.

Third, the actual cost of retrofitting an existing cement plant with oxyfuel or calcium looping technology is uncertain. Especially for oxyfuel, an extended plant shutdown will be required and refurbishing the existing equipment may increase costs relative to purpose-built equipment in a new oxyfuel cement plant. If coordinated with a long shutdown for plant modernization, such costs could be reduced (Hills et al., 2016). Tail-end calcium looping retrofitting will need less plant downtime and modification of the host process than oxyfuel, but successful heat integration and CaO integration could impose additional retrofitting costs. Quantifying these retrofitting costs requires a separate study and is deferred to future work.

Fourth, potential space constraints can present a substantial challenge for retrofitting cement plants (Hills et al., 2016). SARC has a relatively high footprint (more than double that of MEA (Zaabout et al., 2017)), whereas both oxyfuel and calcium looping will also impose substantial, albeit smaller, footprints. A potential mitigating factor for SARC is that positioning the CO_2 capture plant some distance away from the cement plant should not introduce large additional costs because the flue gas stream is the only point of connection. Additional process units for oxyfuel and calcium looping need to be positioned close to the cement plant due to closer integration. If necessary, the SARC footprint could be reduced via additional civil works to position SARC reactors above each other, but this will increase overall capital costs.

Fifth, the development of electricity prices and CO_2 intensities over the plant lifetime is an important uncertainty in determining the cost of CO_2 avoidance for all three technologies. It is safe to assume that CO_2 intensities will reduce over time, but the speed of this decrease and the impact of decarbonization on electricity prices paid by large industrial consumers are uncertain.

Finally, the price and availability of coal for calcium looping could become problematic in developed regions looking to phase out the use of coal. In this case, calcium looping could be made to run on biomass or waste fuels, since fluidized bed combustion is flexible with respect to fuel quality.

It can also be noted that trends in the global energy system towards cleaner electricity are positive for oxyfuel and SARC technologies. In addition, the extreme speed of global decarbonization required to align with the Paris Climate Accord suggests that retrofittability could be of high importance. As these trends push more world regions towards the bottom right of Fig. 14, SARC will become an increasingly attractive alternative.

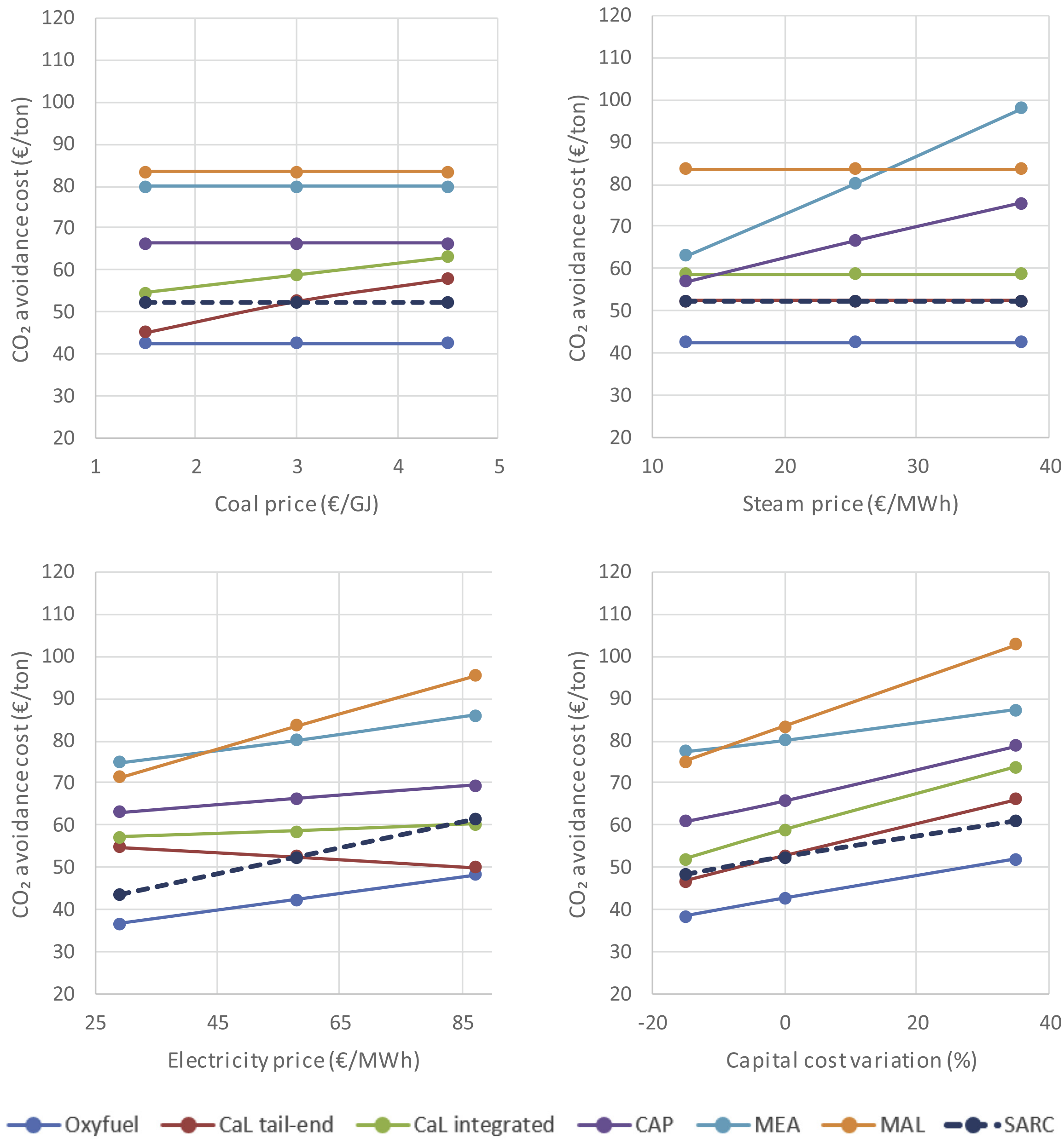

Fig. 12. Economic benchmarking of SARC to other technologies evaluated in the CEMCAP project. CaL = calcium looping; CAP = chilled ammonia process; MEA = monoethanolamine; MAL = membrane-assisted CO_2 liquefaction.

4. Summary and conclusions

The SARC concept is a promising candidate for cost effective CO_2 abatement from the cement industry. This study showed that SARC compares favourably with benchmark technologies in terms of CO_2 avoidance costs and the cost of clinker production.

SARC achieved a CO_2 avoidance cost of €52/ton in the base case, which is about €10/ton higher than oxyfuel technology, almost identical to calcium looping technology and lower than four other alternatives. When more optimistic cost correlations were employed for large-scale vacuum pumps, SARC CO_2 avoidance costs reduced to €45/ton, which is close to the oxyfuel technology. More detailed engineering studies on large-scale vacuum pump costs are therefore recommended.

This is a promising result considering the high retrofittability of the SARC technology. Since SARC only consumes electrical power and requires no integration with the host process, it offers a promising solution to the future scenario of rapidly rising CO_2 prices, where existing industrial processes must rapidly reduce emissions. Other economically attractive CO_2 capture processes such as oxyfuel and calcium looping introduce greater complexity and uncertainty when retrofitting existing plants.

SARC can also benefit economically from two developments that could play out in the future. First, demand for specialized large-scale heat exchange surfaces and vacuum pumps could unlock additional economies of scale beyond those assumed in the base case. Such improved economies of scale can reduce CO_2 avoidance costs by 11%. Second, strong incentives for electrification in industry using renewables could result in lower effective electricity prices with lower carbon intensities. For example, an €18/MWh reduction in electricity price for carbon-free electricity could reduce the CO_2 avoidance cost by 17%.

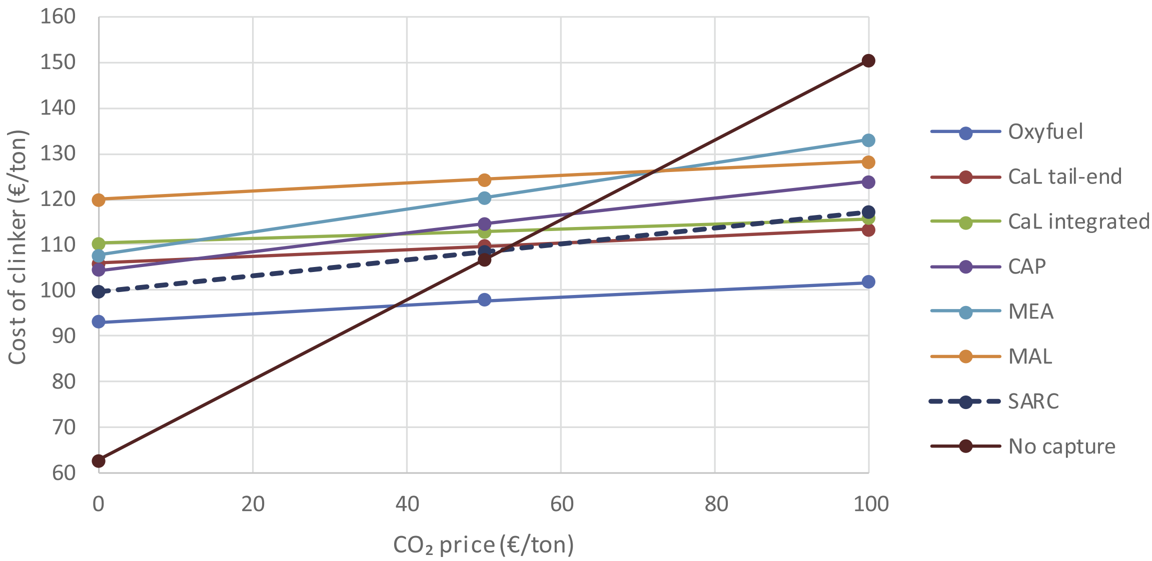

Fig. 13. Sensitivity of different technologies to the CO_2 price.

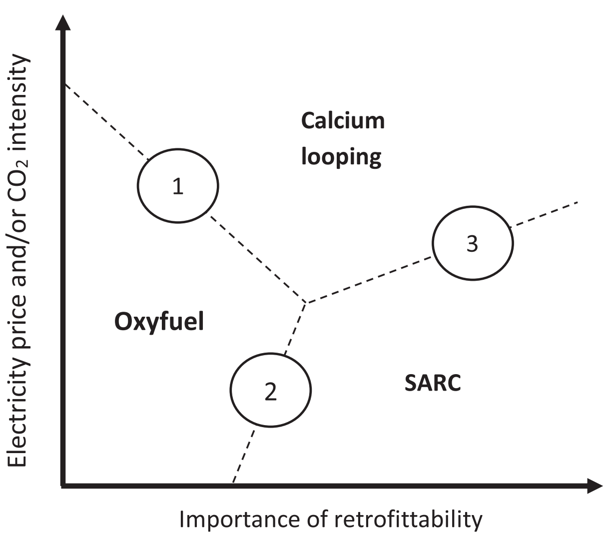

Fig. 14. Qualitative illustration of the conditions under which Oxyfuel, calcium looping and SARC technologies will be most attractive.

SARC is therefore well positioned for the ongoing global energy transition, particularly in scenarios compatible with the Paris Climate Accord that will force existing CO_2 emitters to decarbonize. In general, cleaner electricity makes SARC more attractive relative to calcium looping, while a high demand for retrofitting of existing cement plants makes SARC more attractive relative to oxyfuel. The modular nature and low operating temperature of SARC reactors also promises a faster and more economical demonstration and scale-up process. Further development and scale-up of the technology is therefore recommended.

CRediT authorship contribution statement

Schalk Cloete: Conceptualization, Methodology, Formal analysis, Investigation, Writing - original draft, Writing - review & editing, Visualization, Funding acquisition. **Antonio Giuffrida:** Formal analysis, Investigation, Writing - original draft. **Matteo C. Romano:** Methodology, Writing - review & editing, Supervision, Funding acquisition. **Abdelghafour Zaabout:** Conceptualization, Writing - review & editing, Project administration, Funding acquisition.

Declaration of competing interest

The authors declare that they have no known competing financial interests or personal relationships that could have appeared to influence the work reported in this paper.

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Appendix A. Supplementary data

Supplementary data to this article can be found online at https://doi.org/10.1016/j.jclepro.2020.123024.

Nomenclature

Acronyms

CAC	CO_2 avoidance cost [€/ton]
CAP	Chilled ammonia process
CaL	Calcium looping
COC	Cost of clinker [€/ton]
CSTR	Continuous stirred-tank reactor
EoS	Economies of scale
KPI	Key performance indicator
MAL	Membrane-assisted CO_2 liquefaction
MEA	Monoentanolamine
O&M	Operating and maintenance
PEI	Polyethyleneimine

PFR	Plug flow reactor
SARC	Swing Adsorption Reactor Cluster
TDC	Total direct cost
TPC	Total plant cost

Symbols

C	Levelized cost [€/ton_{clk}]
E	Emissions intensity [ton_{CO2}/ton_{clk}]

Subscripts

cap	Capital
clk	Clinker
dir	Direct
el	Electric
eq	Equivalent
O&M	Operating and maintenance
ref	Reference
RM	Raw meal

Appendix. Cost correlations

The correlations used to calculate the costs of key plant components are summarized below. Costs for compressors, vessels and heat exchangers were estimated using the correlations provided by Turton et al. (2008). Three key equations are used for this purpose.

First, the purchased cost of the equipment assuming carbon steel construction and atmospheric pressure operation (C_p^0) is calculated. The coefficients (K) are given in Table 5 and the scaling parameter (A) is power [kW] for compressors, volume [m^3] for vessels and surface area [m^2] for heat exchangers. In the case of heat exchangers the correlation only applies up to 1000 m^2.

$$\log\left(C_p^0\right) = K_1 + K_2 \log(A) + K_3(\log(A))^2 \quad \text{Equation 4}$$

To find the bare module cost, factors for pressure other than atmospheric and materials other than carbon steel must be specified. Material factors (F_M) are specified in Table 5, while pressure factors (F_P) are calculated using Equation (5) for vessels and Equation (6) for heat exchangers. Here, P is the gauge pressure [bar] and D is the vessel diameter [m]. For a vacuum below 0.5 bar, F_P = 1.25.

$$F_P = \frac{\frac{(P+1)D}{2(850-0.6(P+1))} + 0.00315}{0.0063} \quad \text{Equation 5}$$

$$\log(F_P) = C_1 + C_2 \log(P) + C_3(\log(P))^2 \quad \text{Equation 6}$$

Using the equipment cost, material and pressure factors, the bare module cost (C_{BM}) can be calculated using Equation (7) for compressors and Equation (8) for vessels and heat exchangers. The coefficients (B) are given in Table 5. As discussed in the text, the heat exchanger bare module cost is halved because only the tubes are needed for heat transfer inside the SARC reactors. All these costs are calculated in 2001 \$ with a CEPCI index of 397.

$$C_{BM} = C_p^0 F_M \quad \text{Equation 7}$$

$$C_{BM} = C_p^0 (B_1 + B_2 F_M F_P) \quad \text{Equation 8}$$

Table 5
Coefficients used in the Turton et al. (2008) correlations.

Coefficient	Compressors	Vessels	Heat exchangers
K_1	2.2897	3.4974	4.3247
K_2	1.3604	0.4485	−0.3030
K_3	−0.1027	0.1074	0.1634
C_1	–	–	−0.00164
C_2	–	–	−0.00627
C_3	–	–	0.0123
B_1	–	2.25	1.63
B_2	–	1.82	1.66
F_M carbon steel	2.7	1	1
F_M stainless steel	5.7	3.1	2.7

The vacuum pump bare module cost is calculated according to Equation (9) based on Woods (2007). Here, the reference cost is C_{ref} = 50000, the reference size is S_{ref} = 10, the scaling coefficient is n = 0.64, the installation factor is F_I = 1.7, and the material factor is F_M = 2.1. The material factor was estimated from the compressor material factors in Table 5 as 5.7/2.7. The bare module cost (C_{BM}) is then estimated by specifying the desired volume flowrate in the units of kg/h/kPa, with a maximum value of 500. The correlation returns costs in \$ with a CEPCI value of 1000.

$$C_{BM} = C_{ref}\left(\frac{S}{S_{ref}}\right)^n F_I F_M \quad \text{Equation 9}$$

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