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Techno-economic analysis of a hybrid CO_2 capture system for natural gas combined cycles with selective exhaust gas recirculation

Maria Elena Diego*, Jean-Michel Bellas, Mohamed Pourkashanian

Energy 2050, Faculty of Engineering, University of Sheffield, S10 2TN, United Kingdom

ABSTRACT

This work analyses the implementation of CO_2 capture in natural gas combined cycle (NGCC) power plants using a hybrid system integrated by an amine scrubbing plant and a CO_2 selective membrane. In this configuration, the membrane unit operates at close to atmospheric pressure and it is used to selectively recycle CO_2 back to the inlet of the compressor, therefore increasing the CO_2 content of the flue gas entering the capture system. A novel integration between the amine capture plant and the selective membrane is analysed here, which aims at exploiting the benefits of both the parallel and series selective exhaust gas recirculation (S-EGR) existing options. The mass and energy balances performed on this system indicate that the new configuration generates a flue gas with a CO_2-enhanced concentration of 18%vol., which leads to a decrease in the energy demand in the reboiler by 6% with respect to an amine scrubbing system coupled to a conventional NGCC plant without S-EGR. Moreover, a reduction of 77% is achieved in the gas flowrate fed to the absorber of the amine plant, thus significantly reducing its size and cost. The calculated net electrical efficiency of the plant is 50.3%, which is 0.5 net percentage points higher than that of a conventional NGCC with amine-based capture and slightly lower than that of a reference plant with exhaust gas recirculation (EGR). These values are dependent on the pressure drop associated with the membrane system, which has a large influence on the energy balance of the plant. Therefore, higher efficiency improvements can be achieved if membrane module designs with reduced pressure drop are used. A techno-economic evaluation reveals that the cost of the membrane system has a strong effect on the capital costs of the plant and thus, on the cost of electricity and the cost of CO_2 avoided. These values vary between \$81.9 and \$93.9 per MWh and \$82.6 and \$121.8 per tonne of CO_2 avoided, respectively, for the S-EGR cases studied at a reference capacity factor of 0.85. A sensitivity analysis shows that it is necessary to reduce the costs of the reference hybrid S-EGR system in order to make it competitive against current benchmark options. Therefore, further ongoing development towards membrane units with high CO_2 permeance, limited pressure drop and reduced costs is particularly interesting for the S-EGR system studied in this work. The obtained results also indicate the targeted values of these parameters that can make the cost of the S-EGR configuration to be below that of conventional systems with amine capture and EGR options for CO_2 capture in NGCC power plants under different scenarios.

Keywords: CCS; combined cycles; amine capture; membranes; selective exhaust gas recirculation; EGR

*Corresponding author: m.diegodepaz@sheffield.ac.uk

1. INTRODUCTION

The use of natural gas as a fuel is expected to substantially contribute to the supply of the increasing electricity demand worldwide in the next few decades [1, 2], accounting for 16 to 24% of the total share by 2040 [1]. This makes gas-fired power plants promising candidates for the implementation of CO_2 capture and storage (CCS) technologies in the energy sector (see for example the recent analysis in [3]), which can thus contribute to achieving the stringent targets agreed at COP21 in order to tackle climate change [4]. Nevertheless, the flue gas generated in gas-fired power plants has particular features that make coupling these systems with post-combustion CCS particularly challenging. Natural gas combined cycle (NGCC) power plants operate with high excess air ratios because of gas turbine material constraints, which dictate the maximum allowable inlet temperature. As a result, gas-fired systems generate large flue gas flowrates with a CO_2 concentration as low as ~3-4%vol. These characteristics lead to the need for CO_2 capture reactors with large areas to accommodate the flowrate, which also work under CO_2-restricted driving force conditions thereby increasing the energy required for CO_2 separation [5, 6] and thus, the energy penalties and costs of the associated post-combustion capture plants. This is the case for the amine-based capture systems investigated in this work, where lower reboiler duty and cost savings can be achieved with higher CO_2 concentrations in the exhaust [6-8].

A number of process modifications are being investigated to make the CO_2 capture process in gas-fired power plants more economical by increasing the CO_2 concentration (higher driving force) and reducing the flowrate of the flue gas entering the capture plant [9]. This is the case of natural gas combined cycle configurations that make use of exhaust gas recirculation (EGR) or selective exhaust gas recirculation (S-EGR) schemes, where a fraction of the inlet air is replaced by a chilled recycled gas flow. As a result, a smaller flue gas flowrate with a higher CO_2 concentration is generated [6, 7, 9-19]. In the EGR process, this is achieved by taking a fraction of the flue gas exiting the heat recovery steam generator (HRSG) back to the inlet of the compressor, after passing through a cooling and a water knockout stage. The resulting flue gas after combustion also has a lower O_2 concentration compared to conventional systems without recirculation, which is an additional benefit if amine scrubbing is used as the capture technology because of the potential reduction in the oxidative degradation of the solvent [20]. Nevertheless, associated O_2 levels in the combustor also decrease. Therefore, only modest recirculation ratios can be achieved in these systems to avoid issues related to flame stability and/or combustion efficiency [21-24] that may require major combustor redesign [25]. It is for this reason that many previous works have considered maximum EGR ratios of around 40% [6, 11, 14, 15, 18, 22, 26, 27]. This leads to oxygen levels at the combustor inlet no less than 16%vol., thus preventing undesired combustion effects from happening when using combustors currently available [21, 22, 24, 25]. In these cases, the maximum CO_2 concentration that can be attained in the flue gas is usually limited to around 6.5%vol. [6]. In order to achieve higher CO_2 levels in the flue gas without compromising combustion stability, Merkel et al. [13] suggested the use of selective exhaust gas recirculation. The S-EGR concept relies on selectively recirculating CO_2 back to the inlet of the compressor as shown in Figure 1, thereby avoiding other dilutant gas species present in the flue gas (N_2 mainly). Following this approach, it is possible to increase the CO_2 content in the flue gas far beyond the values of EGR, while keeping sufficiently high O_2 levels in the oxidant stream entering the combustor [13]. In order to separate the CO_2 from the flue gas prior to recycling, Merkel et al. proposed the use of a CO_2 selective membrane. This unit employs an air sweep stream that flows counter-currently to the flue gas and works at close to atmospheric pressure in both the feed/retentate and the permeate streams [13], although other options are also being investigated (i.e., the use of a solid absorbent bed in a rotary wheel as the selective CO_2 separator [14]). The resulting air-flow at the exit of the membrane, which is now rich in CO_2, is then fed to the compressor. These authors indicate that the use of the counter-current feed air stream together with the increased CO_2 content in the flue gas provides sufficient driving force for the CO_2 separation to take place in the selective membrane, without the need for vacuum or compression stages, which are highly energy consuming. Instead, only a slight pressure lift should be provided in both the air and flue gas feed streams to overcome the pressure drop across the membrane system [13]. Therefore, the idea behind the S-EGR concept outlined in Figure 1 is to use the selective membrane only as a CO_2 pre-concentrator, thus increasing the CO_2 content of the flue gas sent to the post-combustion capture system using a limited energy input. This has potential efficiency and economic advantages for NGCC systems provided with post-combustion amine capture systems, as indicated above [9, 13, 19].

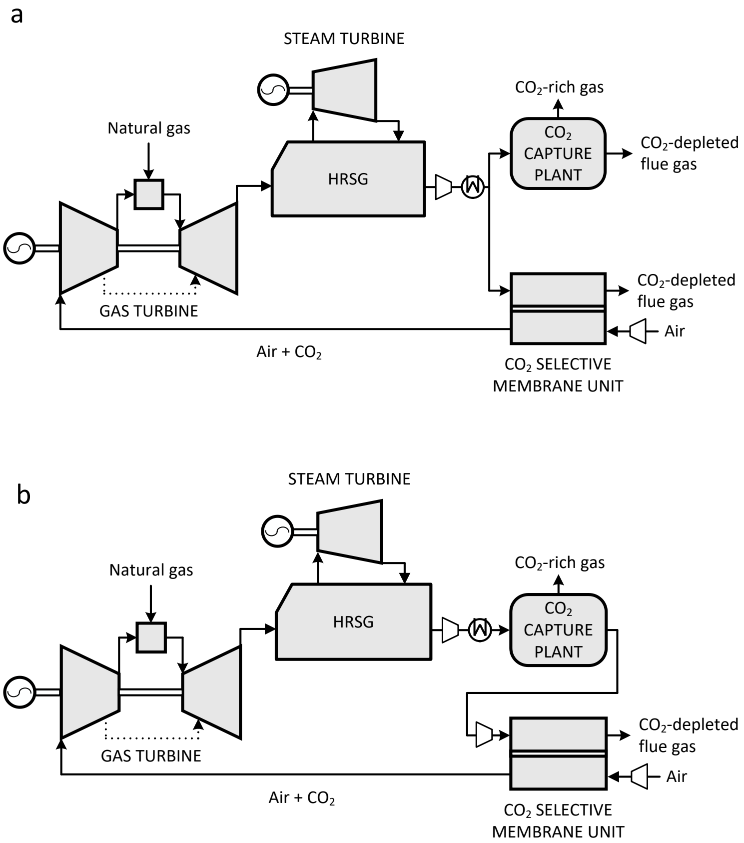

Figure 1. Parallel (a) and series (b) S-EGR schemes for CO_2 capture in NGCC power plants.

Two S-EGR configurations were proposed, which are referred as parallel and series S-EGR (Figures 1a and 1b, respectively) [13]. These can be integrated with a NGCC coupled with an amine scrubbing plant for CO_2 capture, thus constituting a hybrid S-EGR configuration. In the parallel configuration, the flue gas after the HRSG is split into two fractions: one is sent to the amine capture plant whereas the other is passed through the CO_2 selective membrane unit operating counter-currently to the combustion air (see Figure 1a). In the series scheme, the flue gas passes first through the amine capture unit to remove only a fraction of the CO_2, whereas the rest is stripped in the subsequent selective membrane and used for recycling, as shown in Figure 1b. Both configurations have specific benefits and limitations. The size of the amine capture plant and the area of the selective membrane are substantially reduced in the parallel S-EGR system with respect to the series configuration, because of the decreased flue gas flowrate with high CO_2 concentration treated in both units [13-15, 19]. Nevertheless, the CO_2 capture efficiency in the amine system, as well as that in the selective membrane, should be pushed to very high levels (values equal or above 95% are usually considered) to retain a large fraction of the CO_2 generated during combustion as well as all the recirculated CO_2 and attain the required high overall CO_2 capture efficiencies (typically close to 90%) [13-15]. In contrast, the capture plant in the series S-EGR configuration can operate at a much lower capture efficiency (values in between 30 and 58% have been investigated [13, 14]), since the subsequent selective membrane unit strips the remaining CO_2 needed to achieve the overall capture target (also ~90% in these works) [13, 14]. However, in this configuration, all flue gas is sent to the capture plant and is then passed through the membrane (with a reduced CO_2 content) and therefore, large absorber cross-sectional and membrane areas are required.

Only a limited number of works have investigated the potential of coupling NGCC power plants with post-combustion capture systems and the S-EGR processes described above [13-15, 19]. Merkel et al. initially proposed the concept and carried out an analysis of the configurations in Figures 1a and 1b, using Polaris membranes for the selective recycle and a generic CO_2 capture unit [13]. In the parallel S-EGR configuration, these authors fixed the overall CO_2 capture efficiency at 90% and the S-EGR recycle ratio at 77% (ratio between the flue gas sent to the selective membrane and the total flue gas flowrate). Under these conditions, a large increase in the flue gas CO_2 concentration of up to 18.6%vol. was calculated, whilst keeping 16.0%vol. O_2 at the inlet of the combustor. However, this is at the expense of a very high capture efficiency in the CO_2 capture plant as discussed above, which takes a value of 98% in this case, in addition to a CO_2 separation in the membrane of 97% [13]. Merkel et al. also investigated the potential of the series S-EGR configuration. For this purpose, they fixed the CO_2 capture efficiency of the capture unit and that of the selective membrane at around 30 and 96%, respectively, thus attaining close to 91% overall CO_2 capture efficiency. They reported an O_2 content in the oxidant stream fed to the combustor of 16.3%vol. in this case, whereas the flue gas entering the capture unit has 13.7%vol. CO_2 [13]. A detailed S-EGR analysis was also carried out by Herraiz, who investigated cases where the capture unit uses amine scrubbing (30%wt. MEA) as the capture technology, whereas the selective CO_2 separation is carried out by a solid sorbent packed within a rotary wheel device [14]. Three cases were analysed for the series S-EGR case, assuming different combinations of the CO_2 separation efficiency in the amine capture plant and that of the rotary wheel employed as the CO_2 selective separator (i.e., 31/95%, 48/90% and 58/85%, respectively), whilst maintaining the overall system capture efficiency at 90%. In these cases, the CO_2 content in the flue gas was calculated to increase up to 12.9, 8.2 and 6.7%vol., respectively, whereas the O_2 content in the oxidant stream at the inlet of the combustor ranged between 18.9 and 20.2%vol. [14]. For the parallel S-EGR configuration, Herraiz assumed 96% absorption efficiency in the amine capture system together with a CO_2 separation efficiency in the rotary wheel of 97%. A 90% overall capture efficiency was attained under these conditions with a recycle ratio of 70%, thus leading to 14.1%vol. CO_2 in the flue gas and 18.7%vol. O_2 in the oxidant stream [14]. Recently, a techno-economic analysis of the parallel S-EGR scheme has been conducted by Diego et al. [15], using a post-combustion amine capture plant (aqueous MEA 30%wt.) and a CO_2 selective membrane, both working at 95% capture efficiency. The evolution of the overall system capture efficiency and the CO_2 concentration in the flue gas was analysed in this work as a function of the S-EGR recycle ratio. It was calculated that the maximum recycle ratio that could be targeted under these conditions is 53% if an overall CO_2 capture efficiency of 90% is to be achieved [15]. In this case, the CO_2 concentration of the flue gas sent to the capture plant is limited to 8%vol. and the O_2 concentration in the CO_2-enriched air entering the combustor is close to 20%vol. The analysis of this option shows that the economic advantages of this parallel system are largely dependent on the auxiliary consumption and costs associated with the selective membrane skid [15]. In addition to these works that consider hybrid systems for CO_2 scrubbing in the capture plant and the selective separator, some authors have also analysed alternative S-EGR configurations using membranes in series as the only CO_2 separation technology, thus employing compression, vacuum and expansion stages [13, 16]. In these cases, CO_2 concentrations in the flue gas entering the first membrane stage of up to 21.8 [13] and 28.4%vol. [16] have been reported (depending on the O_2 levels at the inlet of the combustor – around 16 and 14%vol., respectively). Also, techno-economic analyses have indicated potential advantages of these configurations against a NGCC coupled with conventional amine capture systems in terms of net electrical efficiency and cost [13, 16]. Similarly, additional S-EGR and EGR combinations for membrane-based systems have been also recently considered [17].

It can be seen from the discussion above that S-EGR can substantially increase the driving force in the CO_2 capture stage, with specific values that depend on the conditions adopted. This is especially interesting for NGCC systems using post-combustion amine scrubbing as the CO_2 capture technology since they could benefit from a relatively mature capture technology with reduced energy penalty as discussed [6, 7]. This is in addition to the use of more compact capture reactors (because of the higher driving force, and also because of the lower flowrates in the parallel scheme) and the expected lower oxidative degradation of the solvent (as a result of the reduced O_2 content in the flue gas). Therefore, S-EGR has the potential to reduce both the capital and the operational costs of the associated amine capture plant systems for NGCCs [9, 14, 15, 17, 19]. The effects on the turbomachinery derived from the new composition of the working fluid in S-EGR schemes are still under discussion, and redesign might be necessary [14, 16, 28]. Moreover, the parallel and series configurations investigated so far also face challenges for its large-scale application. These are specifically the need for very high capture efficiencies in the CO_2 capture plant (parallel configuration), which may be seen difficult to achieve or even unrealistic in certain cases; or for excessive absorber reactor cross-sectional and membrane areas (series configuration). Therefore, it is clear that any improvement in the configurations of Figures 1a and 1b that targets these limitations would be beneficial for the potential application of S-EGR options.

In this context, this work presents and analyses an advanced hybrid S-EGR configuration for CO_2 capture in NGCC power plants that makes use of amine scrubbing in the capture unit and CO_2 selective membranes. This configuration results from a combination of the parallel and series hybrid

S-EGR schemes, thus exploiting the benefits of both configurations and addressing some of the limitations of these systems. Therefore, this study provides an alternative scheme to proposed hybrid S-EGR systems, with a view to its application in large-scale NGCC systems. A techno-economic analysis is carried out to evaluate the potential performance and economic advantages of the new system, focusing on the net electrical efficiency and estimating the associated costs. A benchmark of the proposed configuration against additional CO_2 capture options is also included (NGCC plant using amine capture without and with EGR – namely the ACP and EGR cases, respectively), together with a sensitivity analysis to quantify the effect of key parameters, in order to fully characterise the system. This study provides useful insight on the potential benefits and required development needs of the enhanced hybrid S-EGR configuration presented in this work.

2. PROCESS DESCRIPTION

The proposed hybrid process configuration is shown in Figure 2. This represents a NGCC power plant that makes use of an amine plant for CO_2 capture, in addition to a CO_2 selective membrane (operating at close to atmospheric pressure) to enhance the CO_2 concentration and reduce the flowrate of the flue gas arriving at the capture plant by means of S-EGR. In this new scheme, a fraction of the flue gas that leaves the HRSG is sent to the amine capture plant -which operates at a lower CO_2 capture efficiency than the parallel S-EGR system of Figure 1a-, whereas the remaining flow is directed to a first stage of the CO_2 selective membrane unit that works counter-currently with a CO_2-lean air-flow. As a result, CO_2 permeates through the membrane, thus leading to a CO_2-enriched air stream that is fed to the compressor, and to a flue gas stream partially depleted in CO_2. This flue gas is then mixed with the gas leaving the absorber of the amine system (which still has some CO_2) at a point where both streams have the same CO_2 concentration, thus preventing from CO_2 dilution. This flow, which has a limited CO_2 concentration, is then fed to a second membrane stage, where the inlet air to the gas turbine flows counter-currently as a sweep stream. This stage separates the necessary remaining CO_2 from the flue gas mixture to achieve the desired overall CO_2 capture efficiency, which permeates towards the air-flow used as sweep feed to the first membrane stage. Two gas streams are delivered as a result of the process of Figure 2, namely (i) a CO_2-depleted flue gas that exits the membrane and is emitted into the atmosphere, and (ii) a CO_2-concentrated gas stream which leaves the amine capture plant and can be further purified, compressed and permanently stored.

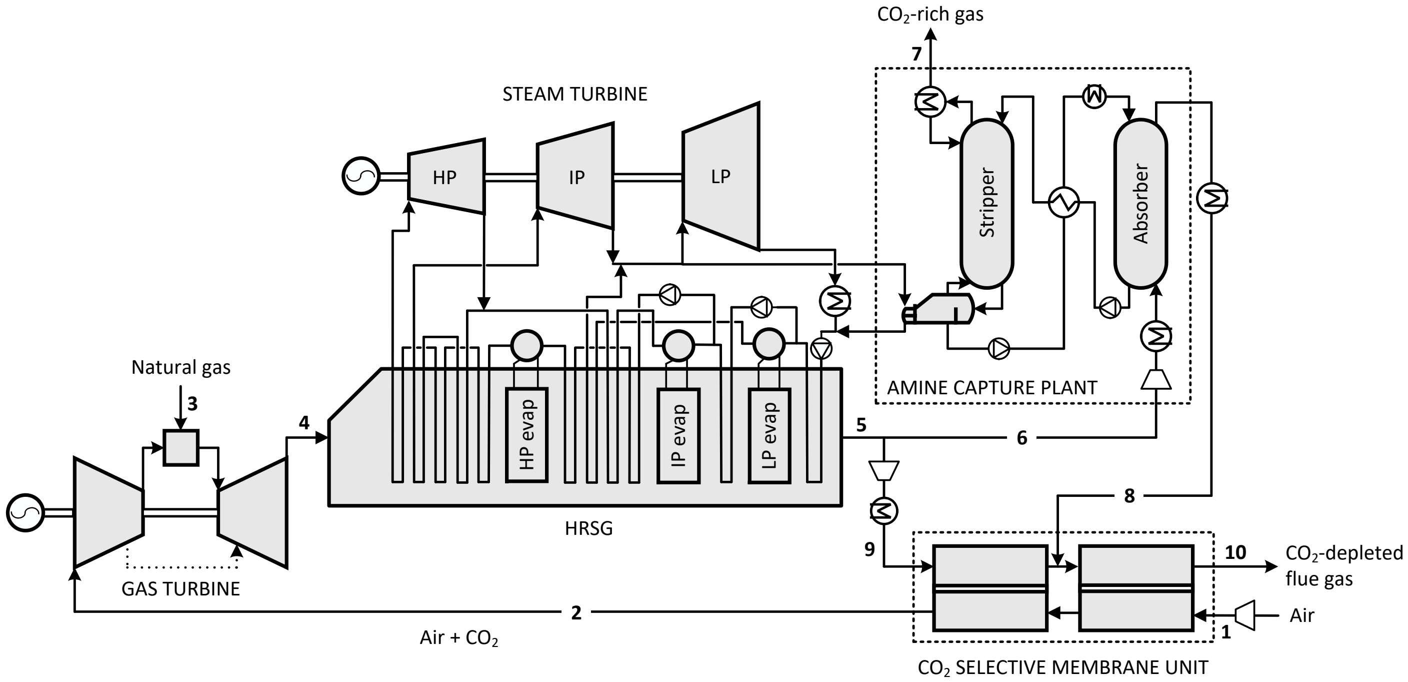

Figure 2. Schematic of the proposed hybrid S-EGR process.

The novel integration between the amine capture plant and the CO_2 selective membrane unit presented here combines the advantages of both the parallel and series hybrid S-EGR configurations mentioned in Figure 1. Similar to the parallel case, only a fraction of the CO_2-enriched flue gas is sent to the amine capture plant, thus reducing the size of the associated CO_2 capture equipment. Furthermore, an important advantage of the system outlined in Figure 2 is that the absorber reactor is not required to work under extremely high CO_2 capture efficiencies (unlike in the parallel configuration), as the exiting flue gas from the amine capture plant is further depleted in CO_2 in the second membrane stage. Moreover, it can be anticipated that the use of this proposed configuration entails important savings in the total membrane area needed with respect to the series configuration under similar conditions, thus leading to savings in the system footprint and cost. This is related to the first membrane stage, which only treats a slipstream of the flue gas with a higher CO_2 concentration (compared to that of the flue gas entering the CO_2 selective membrane after the capture unit in the series scheme of Figure 1b), thus reducing the membrane area requirements necessary to achieve this CO_2 separation. In this context, it is important to highlight that the second membrane stage in Figure 2 is equivalent, in terms of membrane area, to the last part of the membrane system in the series configuration, which strips the remaining CO_2 from the total flow of a CO_2-depleted flue gas to achieve the overall capture efficiency target before it is emitted into the atmosphere (see Figure 1b). Therefore, the overall membrane area needs of the system in Figure 2 reduce when compared to the series configuration.

It is clear from the discussion above that the configuration of Figure 2 could be regarded as an attractive hybrid S-EGR alternative for the integration of CO_2 capture in NGCC power plants. This work analyses the potential of this enhanced scheme, which depends on a balance between the energy and cost reductions achieved in the amine capture plant as a result of S-EGR, as opposed to the energy consumption and cost associated with the CO_2 selective membrane unit.

3. CASE STUDY: METHODOLOGY

A modelling and cost analysis approach is employed to assess the novel S-EGR system proposed. In the following sections, a detailed explanation of the main assumptions and considerations used in this work is outlined.

3.1 Modelling approach

The performance of the configuration outlined in Figure 2 is evaluated using gCCS v1.1.0 (Process Systems Enterprise) [29] as the simulation tool. The modelled NGCC power plant consists of two gas turbines of the GE's 7FA.05 type [25], two HRSGs and a steam turbine (i.e., a 2x2x1 configuration), whose main inlet fuel conditions and operating parameters are summarised in Tables 1 and 2, respectively. In this configuration, the CO_2-enriched air enters the gas turbine system at 30°C after passing through the selective membrane unit, and it is then compressed with a pressure ratio of 17. A total thermal input power of 1103 MW_{th} (LHV) is supplied to the combustor using natural gas (with composition as in Table 1), which resembles that used by DoE/NETL in a reference NGCC power plant without CO_2 capture [25]. The required air inlet flowrate feed into the membrane, which will then enter the combustor together with the recycled CO_2, is calculated to keep the turbine inlet temperature (TIT) at 1275°C. This value is calculated as defined in [30], i.e., as the temperature that results after mixing the combustor outlet flow - at ~1360°C - with all the cooling streams. The pressure drop in the combustor is also assumed to be 5%. After expansion in the turbine, the CO_2-enriched flue gas enters the HRSG, which has three pressure levels with reheat as outlined in Figure 2 (evaporation pressures: 175/28/4 bar (HP/IP/LP); HP superheater/IP reheater outlet temperature: 567°C). The split of the flue gas sent to the amine capture plant and the fraction taken to the selective membrane after the HRSG (which is related to the flow of CO_2 recycled back to the compressor inlet) is adjusted in this configuration to maintain 16%vol. O_2 at the inlet of the combustor, according to the discussion in the Introduction. This limit is selected so that current gas turbine systems could be adapted for the S-EGR conditions considered in this paper without the need for substantial changes to existing combustors [14, 21, 22, 25]. Nevertheless, lower oxygen concentrations could be also potentially achieved (leading to around 2-3%vol O_2 at the combustor outlet) but at the expense of a major combustor redesign [25]. It is important to note that the NGCC power plant model employed in this work has been previously validated in the unabated case against DoE/NETL results [25] using the assumptions in Table 2. Similar assumptions are also used for the simulation of the ACP scheme [15], as well as of the EGR benchmark system, where a fraction of the flue gas is cooled down to 30°C and condensed water is separated before recycling this stream to the inlet of the compressor.

Table 1. Fuel and air composition and properties.

Natural gas	
Inlet temperature (°C)	38
Composition (%mol)	
CH_4	93.10
C_2H_6	3.20
C_3H_8	0.70
C_4H_{10}	0.40
CO_2	1.00
N_2	1.60
LHV (MJ/kg)	47.18
Air	
Ambient temperature (°C)	15
Composition (%mol)	
N_2	77.32
O_2	20.74
H_2O	0.99
CO_2	0.03
Ar	0.92

The operating parameters employed to design the amine capture plant in the S-EGR case and in the additional benchmark CO_2 capture configurations (ACP and EGR schemes) are indicated in Table 2. Two absorber reactors are used to treat the flue gas from the power plant as this configuration can provide greater operational flexibility during part-load scenarios. The temperature of the inlet flue gas stream is reduced to 40°C in a direct contact cooler prior to being fed to the bottom of the absorber, which is assumed to operate at a typical CO_2 capture efficiency of 90%. An aqueous solution of monoethanolamine (MEA 30%wt.) is employed as the solvent, which enters the top of the absorber at 40°C with an optimum lean loading of 0.2 mol CO_2/mol MEA according to previous works [31, 32]. The temperature of the CO_2-rich solvent leaving the absorber is then increased in a heat exchanger before entering the top of the stripper, using the lean solvent exiting the reboiler as the hot fluid and assuming a cold outlet temperature approach of 10°C. In order to supply the necessary heat to the reboiler to allow solvent regeneration in the stripper, a fraction of the low-pressure steam extracted after the intermediate-pressure section of the steam turbine is employed, which is then returned to the steam cycle after the condenser (see Figure 2). The reboiler is taken to operate at 1.75 bar so that the regeneration temperature is below 120°C whilst achieving the desired lean loading, thus avoiding excessive solvent degradation [20, 33]. Moreover, the CO_2-concentrated gas stream that leaves the amine capture plant is taken to a compression and purification unit (CPU) with an estimated energy consumption of 100 kWh/t_{CO2} [34].

Table 2. Summary of the main assumptions used for the simulation of the system in Figure 2.

NGCC power plant	
Thermal input (MW_{th}, LHV)	1103
Temperature of the CO_2-enriched air at compressor inlet (°C)	30
Compressor pressure ratio	17
Compressor isentropic efficiency	83.7
Combustor pressure drop (%)	5
O_2 concentration at combustor inlet (%vol.)	16.0
Gas turbine inlet temperature (°C)	1275*
Gas turbine outlet pressure (bar)	1.05
Gas turbine isentropic efficiency	91.4
Evaporation pressures in HRSG (bar)	
HP	175
IP	28
LP	4
HP Superheater/IP reheater outlet temperature (°C)	567
HRSG pinch point (°C)	10
Steam turbine isentropic efficiency (%)	
HP	88.0
IP	92.4
LP	93.7
Gas pressure drop in HRSG (bar)	0.036
Condenser pressure (bar)	0.048
Pump efficiency (%)	75
Generator efficiency (%)	97
Amine capture plant	
CO_2 capture efficiency (%)	90
Absorber units no.	2
Strippers units no.	1
Solvent lean loading (mol CO_2/mol MEA)	0.2
Gas inlet temperature to absorber (°C)	40
Solvent inlet temperature to absorber (°C)	40
Pressure absorber inlet (bar)	1.14
Column packing	IMTP50
Cold outlet temperature approach rich-lean heat exchanger (°C)	10
Reboiler operating pressure (bar)	1.75
Condenser temperature (°C)	35
Rich/lean pumps outlet pressure (bar)	3
Solvent pumps efficiency (%)	75
Blower efficiency (%)	85
CO_2 counter-current selective membrane	
CO_2 permeance (gpu)	2200
$\alpha_{CO2/N2}$	50
$\alpha_{CO2/O2}$	50
$\alpha_{CO2/Ar}$	50
$\alpha_{CO2/H2O}$	0.7
Pressure drop (%)	5
Blower efficiency (%)	85
Overall system CO_2 capture efficiency (%)	90

*Temperature calculated as that resulting from mixing the combustor outlet stream with the turbine cooling flows, as defined in [30] (combustor outlet temperature ~1360°C).

Taking into account the above information, the absorber and stripper reactors of the amine capture plant are designed following a similar procedure as that described elsewhere [15] and summarised here. The absorber diameter is calculated using typical column design considerations [15, 31, 32], i.e., a flooding factor not higher than 80% and a maximum pressure drop across the reactor of 204 Pa/m (value advised for amine systems, which are moderately foaming [35]), whilst the reactor height is adjusted using 0.1 m steps until the desired capture efficiency is achieved. Similar criteria are followed to estimate the diameter of the stripper reactor, whereas the column height is progressively increased until the reduction in the reboiler duty is deemed negligible (less than 0.05%) after any subsequent increase of 0.1 m in height. This absorber and stripper design procedure is repeated using decreasing liquid-to-gas (L/G) ratios (varied using 0.01 steps), which lead to a reduction in the energy consumption in the reboiler at the expense of a higher absorber height. The optimum L/G ratio is then selected as the value that if reduced further leads to a decrease in the reboiler duty of 0.05% or lower relative to the increase in the absorber height [15]. This methodology allows calculating the absorber and stripper required dimensions, as well as the flow of low-pressure steam that needs to be supplied to the reboiler. The reactor dimensions and values of the reboiler duty obtained using this procedure are in accordance to those reported for systems that operate under similar flue gas conditions [31, 32].

Furthermore, the selective membrane of the hybrid S-EGR system is modelled to account for the transfer of CO_2 and other species (O_2, N_2, H_2O and Ar) between the feed and permeate streams. The permeation of the different gases through the membrane is described in this work using the standard solution-diffusion equation, which provides an effective way to quantify the mass transfer phenomena occurring in the membrane and is suitable for integration with the other process units [16, 36]. This expression correlates the molar flux of each species i through the membrane (J_i) with the permeance (k_i) and the difference in the partial pressure of component i between the feed and the permeate [37]:

$$J_i = \frac{F_i}{A} = k_i\left(P_{feed}x_{i,feed} - P_{permeate}x_{i,permeate}\right) \tag{1}$$

where A is the membrane area and F_i is the molar flowrate of the species i that permeates through the membrane. P_{feed} and $P_{permeate}$ are the pressures of the feed and permeate streams, respectively, whereas $x_{i,feed}$ and $x_{i,permeate}$ represent the molar fraction of the species i in both sides of the membrane.

Furthermore, the membrane selectivity for the component i over the component j (α_{ij}) can be related as the ratio of permeabilities or permeances of these species as indicated in Eq. 2, being the permeance equal to the ratio of the permeability and the membrane thickness [37]:

$$\alpha_{ij} = \frac{k_i}{k_j} \tag{2}$$

A code was developed in gCCS (gPROMS) to simulate the CO_2 selective membrane of Figure 2, which is discretized in the axial direction to describe the variation in the species profile concentration along the system. For this purpose, 20 elements of equal area have been considered to model each membrane stage (similar to the approach in Voleno et al. [36]), which allows for an effective representation of the process with limited computational effort. Permeation of the different species in each of these elements is calculated according to Eq. 1. Then, the calculated output values of each element are employed as the input parameters of the following element. This procedure has been integrated into the membrane code referred above, which can be used to estimate the membrane area required to achieve a CO_2 separation efficiency as well as the flow and composition of the retentate and permeate streams under a set of operating conditions. It is important to highlight that the membrane modelling methodology described here has been successfully validated using data available in the literature [16] before being integrated into the S-EGR process flowsheet (see Table SI-1 in the supplementary information for more details).

The main parameters considered for the membrane simulation in this study are outlined in Table 2. The baseline membrane performance parameters are assumed to be equal to those of the Polaris membrane recently reported by Merkel et al. [13], i.e., CO_2 permeance of 2200 gpu and a CO_2/N_2 selectivity of 50. By using these values together with Eq. 2 it is possible to estimate the N_2 permeance, which is taken to be identical to that of the O_2 and Ar species according to the discussion in DoE/NETL [38]. The H_2O permeance value is then calculated assuming a CO_2/H_2O selectivity of 0.7 [38]. Furthermore, the flue gas entering the membrane system from the HRSG and the absorber is first cooled down to 30°C (knocking out the condensing water). This feed temperature is consistent with the CO_2 permeance and CO_2/N_2 selectivity values mentioned above [13, 39] and is also within the preferred 10-50°C temperature range for this type of membranes (Polaris) [39]. Moreover, different pressure drop values have been reported or assumed for similar S-EGR membrane systems in the literature, ranging from 2.5 to 9% of the inlet pressure [13, 16]. A total pressure drop of 5% is initially considered in this work for the flue gas and air streams passing through the membrane system. Nevertheless, changes in this key parameter are further analysed to account for variations that may arise from different membrane configurations.

3.2 Cost estimation approach

An economic analysis of the hybrid S-EGR configuration of Figure 2 is carried out in this work, assuming the plant is placed in a generic location in the Midwestern US [25] with standard ambient conditions (ISO). In this study, the cost of electricity (COE) and the cost of CO_2 avoided (COA) are used as the economic performance parameters. The COE can be determined as a function of the total overnight costs (TOC), the capital charge factor (CCF), the fixed and variable operating and maintenance costs (FOM and VOM, respectively), the capacity factor of the plant (CF), the net power generated (MW), the net plant heat rate (HR) and the fuel cost (FC). This is given by the following expression [25, 40], where the CO_2 transport and storage costs (T&SC) can be also accounted for:

$$COE = \frac{TOC \cdot CCF + FOM}{CF \cdot 8760 \cdot MW} + VOM + HR \cdot FC + T\&SC \tag{3}$$

The COA is then calculated as the ratio between the increase in the COE and the reduction in the CO_2 emissions rate associated with CCS:

$$COA = \frac{COE_{CCS} - COE_{REF}}{EMS_{REF} - EMS_{CCS}} \tag{4}$$

where COE_{CCS} and EMS_{CCS} are the cost of electricity and the mass CO_2 emission rate of a power plant with CCS, respectively, whereas COE_{REF} and EMS_{REF} are those of a reference power plant without CCS.

The capital cost corresponding to the TOC in Eq. 3 can be estimated as the contribution of the total plant cost (TPC) and other overnight costs (including pre-production, inventory capital, financing, land and other owner's costs) [41]. In turn, the TPC is made up of the bare erected cost (BEC), the engineering, procurement and construction services (EPC), and the costs associated to process and project contingencies [41]. These costs are estimated using a similar procedure as DoE/NETL [25, 41-43], which is detailed below. Moreover, the FOM costs in Eq. 3 include the annual labour for operating and maintenance, administrative and support labour, as well as property taxes and insurance costs. VOM costs account for maintenance material, membrane replacement and consumables costs.

As indicated by DoE/NETL, the bare erected cost comprises the cost of equipment, material and associated direct and indirect labour [25, 42]. In order to account for the differences in scale, the scaled costs (SC) can be calculated for the different equipment areas in Figure 2 using reference plant costs (RC), the corresponding values of the scaling parameter for the reference and the scaled cases (RP and SP, respectively) and scaling exponents (exp) [44], as indicated in Table 3:

$$SC = RC \, . \left(\frac{SP}{RP}\right)^{exp} \tag{5}$$

In this work, the capital costs of a NGCC power plant fitted with CO_2 amine scrubbing reported by DoE/NETL (in 2011 US dollars) are taken as reference (case 1b elsewhere [25]) together with those of the flue gas recycling system (case 1c elsewhere [25]) to calculate the scaled costs of the S-EGR configuration. The capital cost of the gas turbine and the natural gas pipeline and associated components is considered equal to that of the reference plant [25] (case 1b - which has the same thermal input/fuel flowrate, typically used as the scaling parameter) and therefore, these items are not included in Table 3. As discussed above, the minimum O_2 concentration at the combustor inlet chosen for the S-EGR (and EGR) systems is 16%vol. O_2, as this may be feasible with current combustor [22, 24, 25] and gas turbine [14] designs without the need for major modifications. Therefore, this could be expected to have a limited impact on the gas turbine cost for S-EGR applications, although it is still under debate [14, 16, 17, 25]. Moreover, the cost of the membrane system is uncertain at present, as membranes for CO_2 separation (and more specifically, for S-EGR applications) are currently under development and are not yet commercially available for full-scale CCS applications. Nevertheless, a membrane module cost of \$50 per m^2 is initially assumed in this work, since this value has been extensively reported in the literature for membrane systems aimed at CO_2 capture [13, 16, 17, 36, 45, 46].

Table 3. Scaling parameters used in this work.

Specific equipment area	**Scaling parameter**	**Scaling Exponent**
Feed water (FW) system	High pressure FW flowrate	0.72
Steam turbine	Steam turbine power	0.80
HRSG and additional components	HRSG duty	0.70
Cooling water system	Cooling tower duty	0.71
Amine capture plant - absorption	Flue gas flowrate	0.61
Amine capture plant - desorption	Rich solvent flowrate	0.61
Membrane	Membrane area	1
Gas recycling system	Flue gas flowrate	0.70
CO_2 compression and drying	CO_2 flowrate	0.77

Furthermore, the scaling parameters and exponents of the power plant, amine capture plant and CO_2 compression system employed (see Table 3) are selected based on DoE/NETL recommendations [44]. A small variation is made for the case of the amine capture plant, where the scaled costs of the absorber and stripper sections are calculated separately using the flue gas flowrate and the amine rich flowrate as the scaling parameters, respectively, instead of employing the flue gas flowrate only. For this purpose, it is assumed that 65% of the reference amine plant cost corresponds to the absorber reactor and related units (direct contact cooler, blower). The remaining 35% is associated with the desorption side (includes circulation pumps, stripper column, heat exchangers and reboiler), as calculated from the IECM software (version 9.2.1) for a MEA-based CO_2 capture system associated to a NGCC power plant [47]. This allows higher accuracy in the estimation of the capture plant costs because the absorber size (and cost) in the S-EGR case will be smaller compared to the benchmark amine plant used for cost reference due to the reduced flue gas flowrate. Nevertheless, a minor effect is expected on the size (and cost) of the stripper and related components, since the amount of CO_2 to be desorbed will be the same as in the benchmark system (with the same thermal input and overall CO_2 capture efficiency).

Table 4. Main assumptions for the economic analysis.

	Capital charge factor with CCS[a]	0.111
	Capacity factor – CF (%)	85
	Financial cost year	2011
	Plant lifetime (yr)	30
	Fuel cost ($/GJ, HHV)[a]	5.8
	Transport and storage cost (\$/t$_{CO_2}$)[a]	10
TPC & TOC	Cost of installed membrane skid (\$/m^2)[b]	50
	Cost of accessory electric plant, instrumentation and control, improvements to site and buildings and structures (% of process equipment BEC)[c]	20
	EPC cost (% of BEC)[c]	8
	Project contingency (% of TPC)[d]	13
	Process contingency[e]	
	Amine capture plant (% of BEC$_{ACP}$)	20
	CO$_2$ selective membrane (% of BEC$_{mb}$)	20
	Pre-production costs[a]	
	No. of months of all labour	6
	No. of months of maintenance materials costs at 100% CF	1
	No. of months of non-fuel consumables at 100% CF	1
	Percentage of 1 month fuel cost at 100% CF (%)	25
	Miscellaneous (% of TPC)	2
	Inventory capital costs[a]	
	No. of days of consumables at 100% CF	60
	Spare parts (% of TPC)	0.5
	Others[a]	
	Initial cost for chemicals ($/kW)	2.5
	Land costs (M$)	0.3
	Other owner's costs (% of TPC)	15
	Financing costs (% of TPC)	2.7
FOM[a]	Cost of labour ($/h)	51.6
	No. of shifts per day	3
	No. of operators per shift (operating and maintenance)	6.3
	Administrative and support labour costs (% of O&M labour costs)	25
	Taxes and insurance (% of TPC)	2
VOM	Maintenance material cost (% of TPC)[c]	1.1
	Membrane lifetime (yr)[f]	5
	Membrane replacement cost (\$/m^2)[f]	10
	Consumables cost ($/kW)[c]	0.001

[a]Value from assumptions in [25]; [b]Value from references[13, 16, 17, 36, 45, 46]; [c]Calculated from values in [25]; [d]Value from [42]; [e]Value from [41]; [f]Value from [45].

The information above is used to calculate the BEC of the system presented in Figure 2, which also includes the cost of the accessory electric plant, instrumentation and control, improvements to site and buildings and structures. This is calculated as 20% of the BEC associated with the process equipment, as estimated from DoE/NETL costing data for a plant with CO$_2$ capture [25]. Similarly, the engineering, procurement and construction service cost is assumed to be 8% of the total BEC. The techno-economic analysis performed in this paper combines mature elements (e.g. steam turbine, HRSG, etc.) with first-of-a-kind equipment, i.e., the amine capture plant and the CO_2 selective membrane unit. In order to consider this and to account for the cost risk of this novel system due to the lack of operating experience, process and project contingencies were added, according to DoE/NETL recommendations. The project contingency of the NGCC with CO_2 capture is calculated as 13% of the TPC [42]. The process contingency is taken to be zero for the NGCC commercial equipment, whereas it is estimated as 20% of the associated capital cost of the amine capture plant and the CO_2 selective membrane to reflect uncertainties in the cost estimation of these systems [41]. These values are summarised in Table 4, which includes the main assumptions used in the economic analysis. The costing methodology outlined above has been validated using the data from DoE/NETL for a reference NGCC case with an amine capture plant [25], leading to less than a 2% difference in the calculated TOC, COE and COA.

4. RESULTS AND DISCUSSION

The composition and properties of the main streams of the hybrid S-EGR system calculated in the simulation are shown in Table 5. As can be seen, the use of the CO_2 selective membrane allows recycling around 200 kg/s CO_2 from the flue gas to the air stream, thus leading to a gas turbine working fluid with enhanced CO_2 concentration (stream 4). The amount of CO_2 to be recycled is limited by the oxygen concentration at the inlet of the combustor, fixed at 16%vol. O_2 as discussed above. This sets the split of flue gas to the membrane to 76% of the total flowrate after the HRSG, which enters the membrane with 19.5%vol. CO_2 after cooling down to 30°C and condensing out water (stream 9). This flue gas is then mixed with that leaving the absorber at a point where their CO_2 concentration matches, i.e., 2.4%vol. CO_2. Further CO_2 removal is achieved in the last part of the membrane, which together with the first section allows pre-concentrating CO_2 in the oxidant flow up to 14.4%vol. (stream 2). It is calculated that the total membrane area required for the separation of the recycled CO_2 in the S-EGR configuration of Figure 2 is of 2.90 Mm^2, as indicated in Table 6.

Table 5. Composition and properties of the main streams of the hybrid S-EGR configuration.

	Streams									
	1	**2**	**3[a]**	**4**	**5**	**6**	**7**	**8**	**9**	**10**
Mass flow (kg/s)[b]	762	975	23	998	998	237	56	169	726	682
Molar flow (kmol/s)[b]	26.4	31.8	-	33.1	33.1	7.9	1.3	5.9	23.3	23.9
Temperature (°C)	20	30	-	646	111	111	35	30	30	20
Pressure (bar)	1.07	1.01	-	1.05	1.01	1.01	1.70	1.04	1.07	1.01
Composition (%vol.)			-							
N_2	77.3	65.4		62.7	62.7	62.7	0.1	83.2	68.1	85.6
O_2	20.7	16.0		7.1	7.1	7.1	0.0	9.4	7.7	11.4
H_2O	1.0	3.5	-	11.5	11.5	11.5	3.3	4.0	3.9	1.3
CO_2	0.0	14.4		18.0	18.0	18.0	96.6	2.4	19.5	0.6
Ar	0.9	0.8		0.7	0.7	0.7	0.0	1.0	0.8	1.0

[a] Natural gas – see properties in Table 1.

[b] These values correspond to the total flowrates of the S-EGR system (2x2x1 configuration).

Correspondingly, the amine capture plant only treats 24% of the total flue gas at the exit of the HRSG, which has a CO_2 concentration significantly higher than that of conventional NGCC power plants. Table 6 shows a comparison between the S-EGR system analysed in this work and the benchmark ACP and EGR cases for a power plant with the same thermal input (the recirculation ratio in the EGR case is set to 38.6% of the total flue gas –which is recycled at 30°C after cooling and water knockout–, as this gives 16%vol. O_2 at the inlet of the combustor). As can be seen, the flue gas flowrate sent to the amine capture plant reduces substantially in the S-EGR configuration, taking a value of 237 kg/s that corresponds to only 23% and 38% of the total flowrate fed to the amine system in the ACP and EGR cases, respectively. Furthermore, the CO_2 concentration of the flue gas increases up to 18.0%vol. in the S-EGR scheme, compared to the 3.9%vol. CO_2 of the ACP configuration or the 6.4%vol. CO_2 that can be attained in the EGR case. As a result, the design of the amine capture plant is much more compact if S-EGR is used. This is driven by a reduction in the size of the absorber, which has a packing volume 74% and 64% lower than that in the ACP and EGR plants, respectively, as calculated from Table 6. Nevertheless, the stripper design is similar in all the cases since the flowrate of CO_2 absorbed, and thus desorbed, remains the same. This is because all configurations have the same fuel input and capture efficiency, and only some variations in the solvent flow and loading arriving at the stripper occur. Additionally, the higher CO_2 concentration of the flue gas in the S-EGR case also has a positive effect on the solvent flow required in the capture step, which reduces because of the increased driving force in the absorber that favours the CO_2 capture process occurring in the column. This leads to a decrease in the amount of steam required for solvent regeneration and results in a reduced reboiler duty equal to 3.70 MJ/kg CO_2, which is 6% lower than that of a conventional amine capture plant coupled to a NGCC (see Table 6). The differences with respect to the EGR scheme in terms of reboiler duty are more limited. This can be associated to the higher operating temperature of the absorber in the S-EGR case because of the increased CO_2 concentration in the flue gas, thus reducing its performance [10] and limiting the decrease in the solvent flowrate. Moreover, the oxygen concentration at the inlet of the capture system decreases significantly in the S-EGR configuration, especially in comparison with the ACP case (from 12.4%vol. to 7.1%vol. O_2, as outlined in Table 6), which can be beneficial to reduce operating costs in the amine capture plant related to solvent degradation [20].

Table 6. Details of the amine capture plant and selective membrane for the hybrid S-EGR configuration, in comparison with the benchmark ACP and EGR schemes.

	S-EGR	ACP	EGR
Flue gas flowrate to the amine capture plant (kg/s)	237	1030	628
Flue gas composition (%vol.)			
N_2	62.7	74.4	75.4
O_2	7.1	12.4	7.7
H_2O	11.5	8.4	9.6
CO_2	18.0	3.9	6.4
Ar	0.7	0.9	0.9
Amine capture plant design			
Absorber (x2)			
Diameter (m)	7.6	15.0	11.6
Height (m)	17.3	17.1	20.5
Stripper (x1)			
Diameter (m)	7.4	7.6	7.4
Height (m)	27.4	26.9	27.3
LP steam to the reboiler (kg/s)	83.3	89.0	84.4
Reboiler duty (MJ/kg CO_2)	3.70	3.95	3.75
Membrane area (Mm^2)	2.90	-	-
CO_2 emissions after with capture (kg/MWh)	40	41	40

The analysis of the implications of the S-EGR configuration on the plant energy balance is shown in Table 7. The gross power output is equal to 598 MW_e in this system, thus leading to an electrical efficiency of 50.3% after discounting the power required in the CPU and the auxiliaries plant consumption (the water-steam cycle pumps). This is represented in Figure 3, together with the values obtained for the ACP and EGR schemes, as well as those for the case without capture (NGCC w/o CCS) for the sake of comparison. As shown in the figure, the gas turbine power output diminishes in the EGR and most notably, in the S-EGR cases with respect to the ACP and NGCC w/o CCS configurations. This reduction is related to the combined effect of the higher temperature of the CO_2-enriched oxidant stream entering the compressor, which is at 30°C in the S-EGR case and 21°C after mixing with air in the EGR scheme vs. 15°C assumed for the inlet air in the ACP without recirculation and NGCC w/o CCS cases; and the slightly lower calculated mass flowrate of the flue gas entering the turbine (especially in the S-EGR configuration) which also has different composition and properties.

Table 7. Simulation results of the hybrid S-EGR configuration.

Power generation	
Gas turbine power output (MW_e)	396.0
Steam turbine power output (MW_e)	202.3
Total gross power output (MW_e)	598.2
Power consumption	
Power plant auxiliaries (MW_e)	6.3
S-EGR system auxiliaries (MW_e)	11.2
Amine capture plant auxiliaries (MW_e)	5.2
CO_2 compression (MW_e)	20.2
Total power consumption (MW_e)	42.9
Net power output (MW_e)	555.3
Electrical efficiency (%, LHV)	50.3

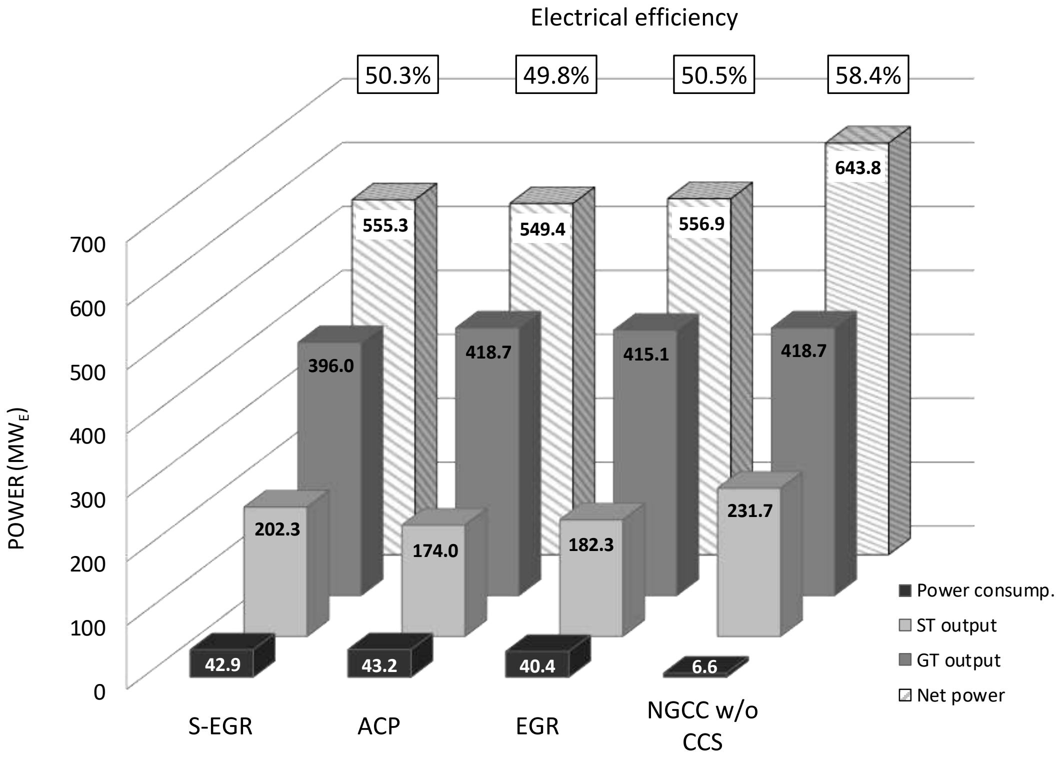

Figure 3. Energy balance of the S-EGR, ACP, EGR and NGCC w/o CCS cases.

In contrast, the power generated in the steam turbine in the S-EGR case is equal to 202.3 MW$_e$, i.e., around 28 and 20 MW$_e$ higher than that in the ACP and EGR schemes, respectively. This is due to the combined effect of the reduced steam extraction required as a result of the lower reboiler duty, and the higher inlet temperature of the flue gas to the HRSG. As a result, the gross power output of the S-EGR system is the highest within the CO_2 capture cases at 598 MW$_e$, followed by the EGR and ACP cases with 597 and 593 MW$_e$, respectively. Moreover, the power consumption for CO_2 compression and purification is the same in all CO_2 capture scenarios, whereas major differences are found in the amine capture plant and the S-EGR/EGR auxiliaries. In the case of the amine capture plant, these are mainly related to the variations in the power required by the flue gas blower and are therefore lower for the S-EGR system (5 MW$_e$) than for the EGR (12 MW$_e$) and the ACP (18 MW$_e$) plants. Nevertheless, further power consumption comes from the need to blow the flue gas and air streams in the S-EGR case before entering the membrane unit to overcome the pressure drop in this device, and from the recycle flue gas fan in the EGR scheme. This leads to a similar total auxiliary consumption in all capture cases (see black bars in Figure 3). Overall, the results indicate that the electrical efficiency of the S-EGR configuration is 0.5 net percentage points higher than that of the ACP case. This leads to a S-EGR efficiency penalty of 8.1 net percentage points when compared to the NGCC power plant without capture, vs the 8.6 net percentage points penalty associated to the ACP case. This makes S-EGR competitive against the conventional capture option for NGCCs in terms of efficiency, although it is slightly lower than that in the EGR configuration.

4.1 Sensitivity analysis

Membrane systems for CO_2 separation and more specifically, for S-EGR applications, are currently under development and are not yet commercially available for full-scale CCS applications. Therefore, there are still uncertainties in the values of key parameters, which can be optimised with further advancements in this technology. This is the case of the pressure drop and the CO_2 permeance of the membrane unit, whose effects are analysed below.

4.1.1 Effect of the pressure drop across the CO_2 selective membrane unit

The pressure drop in the membrane system affects the energy balance of the S-EGR plant, as it is linked to the auxiliary energy consumption [39]. This is shown in Figure 4, which depicts the values of the auxiliary power consumption associated with the hybrid S-EGR plant when the total pressure drop across the membrane unit is equal to 2.5, 5 and 10% of the inlet pressure for both the flue gas and the air streams. As outlined in the figure, an increase in the membrane pressure drop leads to a higher energy consumption in the blowers of the S-EGR system, whereas the auxiliary consumption in the power plant, amine capture system and the CPU is similar in all cases. As a result, the net electrical efficiency of the S-EGR power plant varies from 50.8 to 49.5% for a pressure drop of 2.5 and 10%, respectively (see Figure 4). This indicates that the hybrid S-EGR configuration can be competitive against both ACP and EGR in terms of net electrical efficiency if membrane units with low associated pressure drops are used.

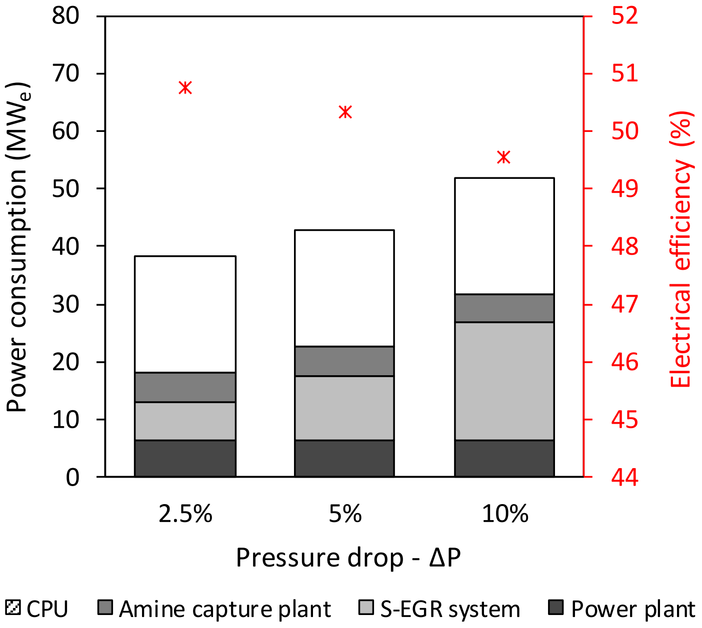

Figure 4. Effect of the pressure drop across the membrane unit in the S-EGR auxiliary consumption and the electrical efficiency of the power plant.

Moreover, the total membrane area required is also slightly affected by these changes. Higher values of pressure drop in the membrane unit require a higher pressure of the flue gas at the membrane inlet, which favours CO_2 permeation towards the air-flow at the beginning of the membrane system.

This leads to limited variations of the membrane area for the cases studied in this work, thus taking values of 3.01 and 2.76 Mm^2 when the pressure drop is 2.5 and 10%, respectively. Nevertheless, the slight reduction in the membrane area requirements for systems with increased pressure drop is at the expense of a higher energy penalty as discussed above. Therefore, the use of membrane module designs with reduced pressure drop is essential to increase the competitiveness of S-EGR systems.

4.1.2 Effect of the membrane CO_2 permeance

The membrane CO_2 permeance is related to the total area required for the selective separation, which decreases with increasing CO_2 permeance values. This is especially important for the S-EGR process in Figure 2, where the CO_2 selective membrane operates at close to atmospheric pressure and thus, large areas are required. In order to illustrate this effect, CO_2 permeances between 1000 and 10000 gpu were considered, as shown in Figure 5.

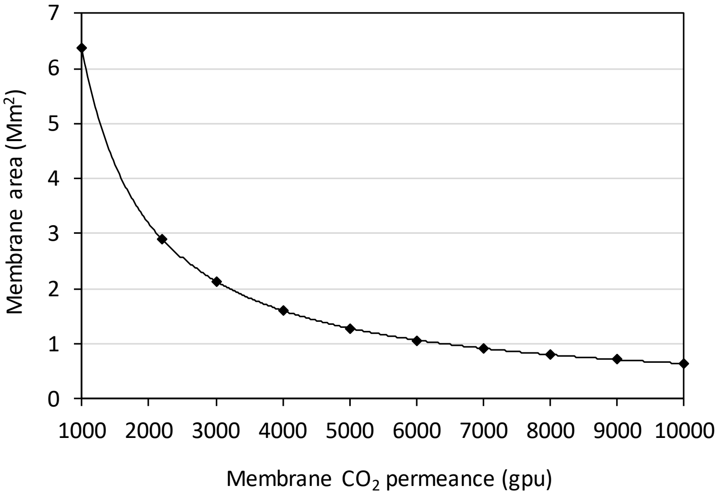

Figure 5. Evolution of the total membrane area as a function of the membrane CO_2 permeance.

As can be seen in the figure, the CO_2 permeance has a direct effect on the total membrane area requirements, which is especially pronounced for low values of this parameter (below 3000 gpu). Thereby, sharp reductions could be achieved in the membrane area if future membranes with higher CO_2 permeance were developed (with similar CO_2 selectivities). For example, using a membrane with a CO_2 permeance of 5000 gpu could reduce the total membrane requirements up to 56% compared to the 2200 gpu baseline case, and if CO_2 permeances of 10000 gpu were achieved this figure would go up to 78%. Membranes for S-EGR applications with increasing CO_2 permeance have been developed in a few years, starting from 1000 gpu and recently getting up to 3000 gpu (laboratory scale) with similar CO_2/N_2 selectivity [39]. Therefore, rapid progress is being carried out in this field [17, 39] and any further advancements can make S-EGR configurations more competitive, significantly decreasing the cost and the footprint of these systems.

5. ECONOMIC ANALYSIS

The information obtained above is integrated within the economic evaluation to estimate the performance of the hybrid S-EGR configuration in terms of cost of electricity and of CO_2 avoided. The results are shown in Table 8 for the S-EGR case at three different membrane CO_2 permeances (2200, 5000 and 10000 gpu) and baseline pressure drop (5%), in comparison with the ACP and EGR systems. As can be seen in this table, incorporating S-EGR substantially decreases the capital costs associated with the amine capture plant. This is a result of the reduction in the absorber size because of the lower flue gas flowrate to be treated, which leads to a drop in the amine plant cost of up to 40 and 27% compared to the ACP and EGR options, respectively. Nevertheless, the need for large membrane areas and the associated S-EGR equipment (i.e., blowers, gas cooler) add to the capital cost of the power plant and lead to a higher TOC than the ACP and EGR schemes for the S-EGR cases considered in Table 3, although differences reduce with increasing membrane CO_2 permeances. This highlights the large effect that any improvement in the membrane CO_2 permeance has on the capital expenditure (CAPEX) of the power plant. Increasing the CO_2 permeance from 2200 to 5000 gpu reduces the costs associated with the membrane system by 56% (proportional to the reduction in area as indicated in Section 4.1.2) and the TOC of the plant by 14%. Similarly, if a membrane with 10000 gpu CO_2 permeance were used, the TOC would be 20% lower than that of the baseline case (2200 gpu), thereby making S-EGR systems more competitive.

Table 8. Capital expenditure costs of the S-EGR, ACP and EGR cases.

	Unit		**S-EGR**		**ACP**	**EGR**
Membrane CO_2 permeance	gpu	2200	5000	10000		
Equipment area	M$					
FW unit and natural gas pipeline		47.9	47.9	47.9	44.7	45.3
Gas turbine system		112.4	112.4	112.4	112.4	112.4
Steam turbine system		59.6	59.6	59.6	52.9	54.9
HRSG system		48.7	48.7	48.7	45.8	46.5
Cooling water system		23.6	23.6	23.6	18.4	20.4
Amine capture plant		105.5	105.5	105.5	176.7	144.3
Membrane system		145.2	63.9	31.9	-	-
S-EGR equipment		31.0	31.0	31.0	-	
EGR equipment		-	-	-	-	16.5
CO_2 compression and drying		28.4	28.4	28.4	28.4	28.4
Accessory electric plant etc.		120.4	104.2	97.8	95.8	93.7
BEC	M$	722.7	625.1	586.8	575.0	562.3
EPC	M$	57.8	50.0	46.9	46.0	45.0
Project and process contingency	M$	174.3	139.8	126.3	133.4	123.9
TPC	M$	954.7	814.9	760.0	754.4	731.2
TOC	**M$**	**1161.4**	**993.2**	**927.2**	**920.4**	**892.5**

The calculated COE and COA values for the configurations in Table 8 are presented in Figure 6. As expected, the main contributor to the COE is the cost of the fuel, whereas the TOC cost also has a key effect. This is related to the equipment costs discussed above, which also have a marked influence on the operating and maintenance costs (the VOM costs mainly) of the plant (see Table 4). The combined FOM and VOM costs are M$50, M$43 and M$40 per year for the S-EGR system with a CO_2 permeance of 2200, 5000 and 10000 gpu, respectively, and they are equal to M$39 and M$38 per year in the ACP and EGR cases, respectively. As a result, the calculated COE of the baseline S-EGR system (2200 gpu) is equal to \$93.0/MWh, which is higher than that of the ACP and EGR cases considered here, at a value of \$84.7 and \$82.7 per MWh, respectively. The COE of the S-EGR system significantly reduces to \$86.8/MWh if membranes with a CO_2 permeance of 5000 gpu are used, and this decreases further to \$84.3/MWh in the 10000 gpu case, which is slightly lower than that of the ACP scheme. A similar trend is followed by the COA, which has a value of \$118.7, \$98.5 and \$90.6 per tonne of CO_2 avoided for the S-EGR cases at a CO_2 permeance of 2200, 5000 and 10000 gpu, respectively, as outlined in Figure 6b. This shows the large influence of membrane properties on the costs of the S-EGR option and confirms that future developments focused on increasing the membrane CO_2 permeance are key for the hybrid S-EGR system of Figure 2.

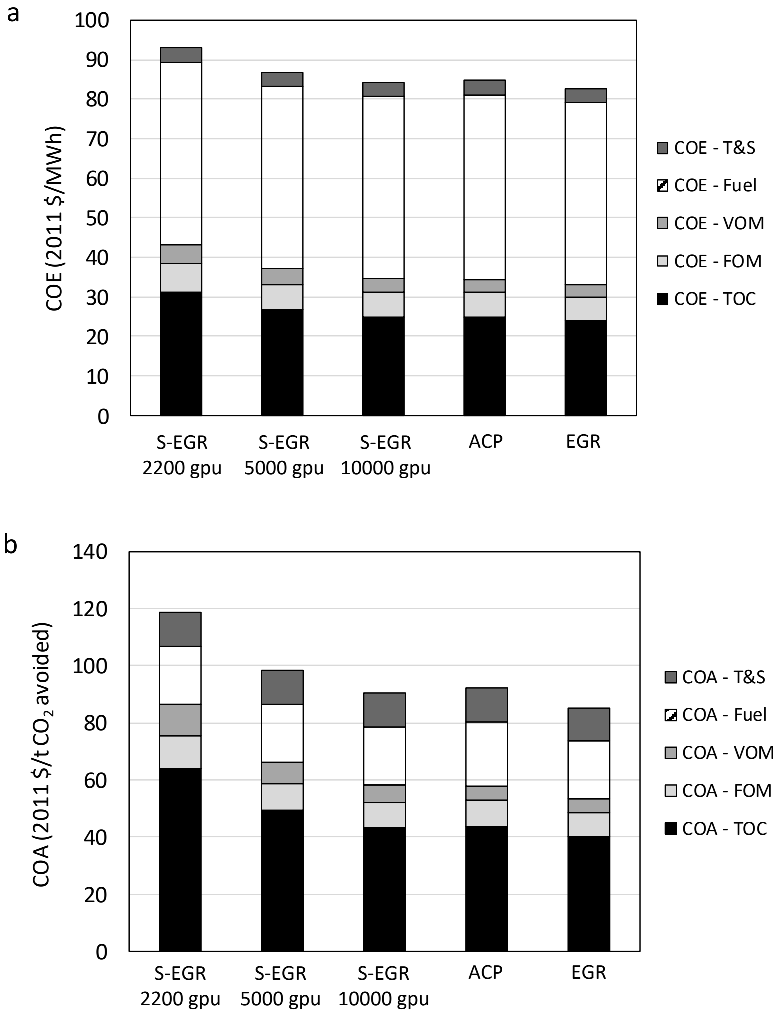

Figure 6. COE (a) and COA (b) for the S-EGR configuration with membrane CO_2 permeance of 2200, 5000 and 10000 gpu and pressure drop of 5%, compared to the ACP and EGR schemes.

It is also important to point out that membranes for CO_2 capture, and specifically for S-EGR applications, are currently under development with large efforts being devoted to research and progress of these systems. This means that large uncertainties can be expected in some parameters used to estimate the costs of S-EGR schemes, which could vary if enhanced performance and cost reductions are going forward. This is the case of the membrane installed skid cost, which is assumed to be equal to \$50 per m^2 in this work, according to the available literature [13, 15-17, 36, 45, 46]. This value is commonly associated with membrane units operating under pressure/vacuum conditions, and it could be expected that future developments (especially for membranes operating at close to atmospheric pressure) could bring these costs down. In order to illustrate the effects of any future reductions in this cost, the COE and COA of the hybrid S-EGR configuration were estimated assuming values between \$10 and \$50 per m^2 for systems with different membrane CO_2 permeance values. These results are outlined in Figure 7, which shows that large reductions in the COE and COA can be achieved if ongoing R&D efforts bring membrane modules costs down.

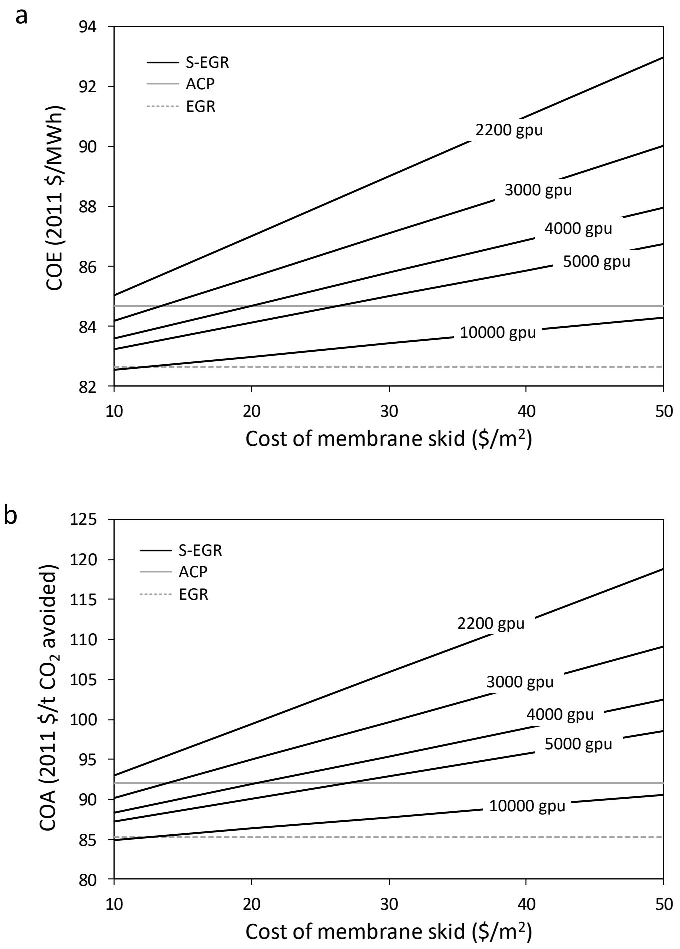

Figure 7. Influence of the installed membrane skid cost on the COE (a) and COA (b) of the hybrid S-EGR configuration for different membrane CO_2 permeances.

Figure 7 indicates that the COE and COA of the hybrid S-EGR system with a CO_2 permeance of 10000 gpu is always lower than that of the ACP case for the membrane price range considered. Larger cost reductions should, however, be targeted in the membrane skid to achieve similar cost values to the EGR scheme. Moreover, the S-EGR configuration could become competitive against the ACP case if the cost of the installed membrane skid is reduced to \$27 per m^2 using a membrane with a CO_2 permeance of 5000 gpu, thus reducing the COE and COA to around \$85/MWh and \$92/t CO_2

avoided, respectively. Further cost reductions would be required if membranes with a lower CO_2 permeance are employed.

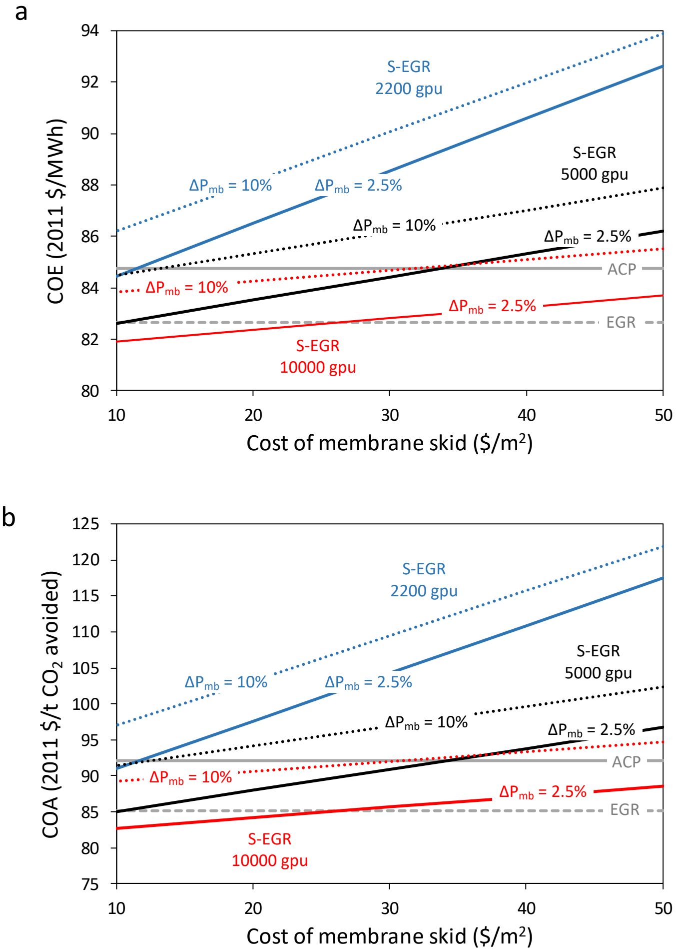

Figure 8. Evolution of the COE (a) and COA (b) of the S-EGR system for different membrane CO_2 permeances, pressure drop and installed skid costs.

Nevertheless, these values also depend on the pressure drop across the membrane unit, which affects the energy balance of the plant as discussed in section 4.1.1. This is shown in Figure 8, where the COE and COA of the S-EGR system are calculated as a function of the cost of the installed membrane skid for different membrane CO_2 permeances (2200, 5000 and 10000 gpu) and pressure drops ranging between 2.5 and 10% of the inlet pressure. As shown in this figure, using a membrane system with the same CO_2 permeance and cost but different pressure drop can vary the COE and COA of the plant by around \$2/MWh and \$6/t CO_2 avoided, respectively, for the conditions studied. This significantly changes the cost reduction targets for the membrane skid, thus affecting the competitiveness of the hybrid S-EGR scheme. For example, the S-EGR scheme using a membrane with a CO_2 permeance of 5000 gpu would be economically competitive against the ACP configuration at an installed membrane skid cost of \$34/m^2 if the system pressure drop is 2.5%, whereas this cost needs to be reduced up to \$13/m^2 for 10% pressure drop.

Finally, the effect of the capacity factor on the COE and COA is also analysed. Gas-fired power plants are expected to operate at varying annual capacity factors in order to match energy demand and supply whilst accommodating the increasing intermittent renewables capacity [48]. This, together with the addition of CO_2 capture systems, will have an impact on costs as shown in Figure 9. This figure depicts the COE and COA of the S-EGR system, compared to the EGR and ACP cases. Various membrane CO_2 permeances (2200 gpu, 5000 gpu and 10000 gpu) and a reference pressure drop of 5% are considered, together with membrane costs of \$50 and \$10 per m^2 and capacity factors ranging from 0.6 to 0.9. As expected, reducing the capacity factor increases significantly the COE and COA of the plant for all cases. This is a result of the effective decrease in the number of annual operating hours, which affects the cost contribution of the capital and fixed operating costs to the COE and COA (see equations 3 and 4) [48]. Therefore, a lower capacity factor means that the number of hours over which the CAPEX can be recovered reduces, whilst the variable operating costs decrease accordingly. Consequently, the higher the CAPEX, the higher the impact of reduced capacity factors [48].

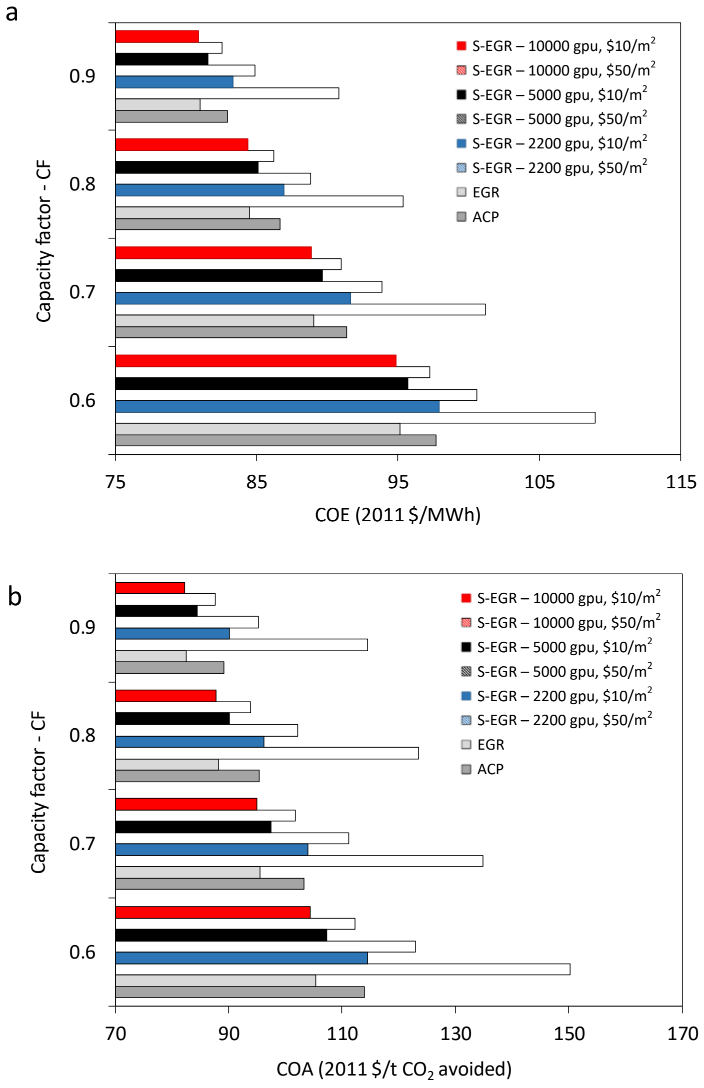

Figure 9. Variation of the COE (a) and COA (b) of the S-EGR, EGR and ACP cases as a function of the capacity factor.

This effect can be seen in Figure 9a by comparing the variations in the COE (and COA, Figure 9b) for an S-EGR system that uses a membrane unit with the same CO_2 permeance but a membrane skid cost of \$50 and \$10 per m^2. The cases with a higher cost of the membrane skid have a higher TOC and, therefore, a decrease in the capacity factor increases the COE and COA in a larger extent. For example, for an S-EGR system which uses a selective membrane with a CO_2 permeance of 2200 gpu, reducing the capacity factor from 0.9 to 0.6 implies an increase in the COE of 19.9% for a membrane skid cost of \$50/m^2 (from \$90.9 to \$109.0/MWh), whereas it is 17.6% for the \$10/m^2 case (varying from \$83.3 to \$97.9/MWh). Similar trends are followed in the 5000 gpu and 10000 gpu options, where the COE and COA increase with lower capacity factors is less pronounced due to the reduced TOC of the configurations (see Table 8). Furthermore, a change in the capacity factor also affects the benchmark ACP and EGR configurations, leading to analogous costing trends as those discussed above for a fixed capacity factor of 0.85. Therefore, an S-EGR system that uses a membrane unit with CO_2 permeance of 5000 gpu and 5% pressure drop can become competitive with the ACP configuration at a cost of the membrane skid slightly below \$30/m^2 for the capacity factors analysed. Moreover, the COE and COA of the S-EGR system is always lower than that of the ACP case for a CO_2 membrane permeance of 10000 gpu (and pressure drop equal to 5%) for the ranges considered.

The hybrid S-EGR configuration proposed in this study uses a novel integration between the amine capture plant and the CO_2 selective membrane unit. This can address some of the limitations of the existing hybrid parallel and series S-EGR systems, namely the need for very high efficiencies in the amine capture plant (characteristic of the parallel S-EGR configuration) or for very large membrane areas (associated to series systems), which affect the system footprint and cost. The techno-economic analysis conducted for a hybrid S-EGR system coupled to a large-scale NGCC expands the database of CO_2 capture options for gas-fired power plants and provides interesting data and trends that can be used to further advance on the future application of gas CCS. Furthermore, the economic results discussed above indicate that developing membrane systems with increased CO_2 permeance, reduced pressure drop and cost is key for the hybrid S-EGR configuration analysed here, which could be competitive against conventional amine capture systems for CO_2 capture in NGCC power plants, in various scenarios. Additional advantages with regards to EGR systems will depend on further optimisation of S-EGR schemes. This could involve the use of improved absorber designs to reduce the temperature in the absorber (e.g., intercooling), which shows a sharper temperature profile in the S-EGR case related to the higher CO_2 content in the flue gas [10]. Therefore, any reduction in the absorber temperature would be predominantly beneficial for S-EGR, thus achieving additional reductions in the energy consumption in the amine capture plant and improving the performance and economics of the overall system against ACP and EGR schemes. Furthermore, a future trade-off between the targeted CO_2 content increase in the flue gas entering the absorber and the associated cost of the membrane system for the separation of the recirculated CO_2 may be advantageous to optimise the cost of the S-EGR configuration and advance in the potential application of hybrid S-EGR systems for CO_2 capture in NGCC power plants. The results obtained could provide a pathway for future R&D investigations, thus indicating the targeted advancements required for hybrid S-EGR systems to be competitive and progress on the implementation of CO_2 capture in NGCCs at large scales of operation.

6. CONCLUSIONS

A hybrid S-EGR configuration using a novel integration between an amine capture system and a CO_2 selective membrane for CO_2 capture in NGCC power plants has been analysed. This system exploits the advantages of both the parallel and series S-EGR schemes, using the membrane unit as CO_2 pre-concentrator and amine scrubbing as the capture technology. The results show that the S-EGR configuration proposed in this study results in a flue gas with a CO_2-enhanced concentration of 18%vol. and reduces the flowrate to be treated in the amine capture plant by 77% and 62% with respect to a conventional ACP reference system without S-EGR and to an EGR scheme, respectively. These features allow a more compact design of the amine capture plant, thus reducing the capital costs of this equipment area by 40 and 27% compared to the ACP and EGR cases, respectively. Additionally, the reboiler duty in the S-EGR case reduces to 3.70 MJ/kg CO_2, which is 6% lower than the ACP case. The calculated net plant efficiency of the baseline S-EGR configuration is 50.3%, which is highly affected by the auxiliary consumption associated with the CO_2 selective membrane. A sensitivity analysis shows that module designs aiming at reducing the pressure drop across the membrane system from 5 to 2.5% can lead to S-EGR net electrical efficiencies of 50.8%, thus being competitive against both the ACP and EGR configurations.

The results show that the final economic benefits of the S-EGR option depend on the balance between the reduced amine capture plant costs and the net plant efficiency gain as opposed to the increase in the capital and O&M costs linked to the membrane unit. This was evaluated in several scenarios varying the membrane CO_2 permeance, installed skid cost and pressure drop across the unit to account for the influence of the membrane in the COE and COA of the S-EGR scheme. The results indicate the hybrid S-EGR scheme costs can vary between \$81.9-93.9/MWh and \$82.7-121.9/$t_{CO2\ avoided}$ for the cases considered (at a reference capacity factor of 0.85). The calculated costs of the S-EGR baseline case (CO_2 membrane permeance of 2200 gpu, selectivity of 50) are higher than those of the benchmark processes due to the associated membrane costs. Nevertheless, a sensitivity analysis has shown that the hybrid S-EGR configuration can be competitive against the ACP case in terms of COE and COA in various scenarios. This requires to continue further ongoing development on membrane technology in order to advance on membrane systems with increased CO_2 permeance, reduced costs and pressure drop, which are key for the competitiveness of the hybrid S-EGR system. Advantages with respect to the EGR case could be also achieved depending on the advancements in membrane technology and optimisation of the S-EGR system. The analysis carried out in this study highlights the potential benefits and remaining development needs of hybrid S-EGR systems, and expands the database of CO_2 capture options for gas-fired power plants.

ACKNOWLEDGEMENTS

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SUPPLEMENTARY INFORMATION

Table SI-1. Inputs and calculated values for the validation of the membrane model used in this work.

Parameter	**Turi et al. [16]**	**This work**	**Turi et al. [16]**	**This work**
	Input parameters (from Turi et al. [16])			
Membrane properties				
Membrane CO_2 permeance (gpu)	3500		1000	
$\alpha_{CO2/N2}$	35		50	
$\alpha_{CO2/O2}$	35		5	
$\alpha_{CO2/Ar}$	35		5	
$\alpha_{CO2/H2O}$	0.7		0.3	
CO_2 separation efficiency (%)	97.9		97.7	
Inlet streams properties				
Inlet flue gas				
Flowrate (kg/s)	575		578	
Pressure (bar)	1.98		1.98	
Composition (%vol.)				
N_2	63.9		66.5	
O_2	5.5		5.4	
CO_2	25.2		22.5	
H_2O	4.8		4.9	
Ar	0.7		0.8	
Pressure drop in the feed side of the membrane (%)	2.5		2.5	
Inlet air				
Flowrate (kg/s)	394		430	
Pressure (bar)	1.07		1.07	
Composition (%vol.)				
N_2	77.3		77.3	
O_2	20.7		20.7	
CO_2	0.0		0.0	
H_2O	1.0		1.0	
Ar	0.9		0.9	
Pressure drop in the permeate side of the membrane (%)	5		5	
	Output values			
	From Turi et al. [16]	**Calculated values**	**From Turi et al.[16]**	**Calculated values**
Outlet flue gas				
Flowrate (kg/s)	340	337	376	372
Pressure (bar)	1.93	1.93	1.93	1.93
Composition (%vol.)				
N_2	88.6	88.9	87.7	88.1
O_2	8.9	8.8	10.0	9.9
CO_2	0.8	0.8	0.7	0.7
H_2O	0.6	0.6	0.5	0.6
Ar	1.1	1.0	1.1	0.7
Outlet CO_2-enriched air				
Flowrate (kg/s)	629	631	633	636
Pressure (bar)	1.01	1.01	1.01	1.01
Composition (%vol.)				
N_2	58.3	58.3	60.6	60.5
O_2	13.9	14.0	13.7	13.8
CO_2	22.5	22.4	20.1	20.0
H_2O	4.7	4.7	4.9	4.8
Ar	0.6	0.7	0.7	0.9
Total membrane area (Mm^2)	0.37	0.37	1.34	1.32

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