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Process Analysis of Selective Exhaust Gas Recirculation for CO_2 Capture in Natural Gas Combined Cycle Power Plants using Amines

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ABSTRACT

Post-combustion CO_2 capture from natural gas combined cycle (NGCC) power plants is challenging due to the large flow of flue gas with low CO_2 content (~3-4%vol.) that needs to be processed in the capture stage. A number of alternatives have been proposed to solve this issue and reduce the costs of the associated CO_2 capture plant. This work focuses on the selective exhaust gas recirculation (S-EGR) configuration, which uses a membrane to selectively recirculate CO_2 back to the inlet of the compressor of the turbine, thereby greatly increasing the CO_2 content of the flue gas sent to the capture system. For this purpose, a parallel S-EGR NGCC system (53% S-EGR ratio) coupled to an amine capture plant using MEA 30%wt. was simulated using gCCS (gPROMS). It was benchmarked against an unabated NGCC system, a conventional NGCC coupled with an amine capture plant (NGCC+CCS), and an EGR NGCC power plant (39% EGR ratio) using amine scrubbing as the downstream capture technology. The results obtained indicate that the net power efficiency of the parallel S-EGR system can be up to 49.3% depending on the specific consumption of the auxiliary S-EGR systems, compared to the 49.0% and 49.8% values obtained for the NGCC+CCS and EGR systems, respectively. A preliminary economic study was also carried out to quantify the potential of the parallel S-EGR configuration. This high-level analysis shows that the cost of electricity for the parallel S-EGR system varies from 82.1-90.0 \$/MWh$_e$ for the scenarios considered, with the cost of CO_2 avoided being in the range of 79.7-105.1 \$/ton CO_2. The results obtained indicate that there are potential advantages of the parallel S-EGR system in comparison to the NGCC+CCS configuration in

some scenarios. However, further benefits with respect to the EGR configuration will depend on future advancements and cost reductions achieved on membrane-based systems.

INTRODUCTION

Ongoing commitments to reduce CO_2 emissions and move towards energy decarbonization have prompted the power generation industry to consider less carbon intensive fuels to meet the increasing electricity demand. This is the case for natural gas, where consumption is expected to rise 77% by 2040, accounting for 23% of the global energy mix [1]. However, unabated gas-fired power plants still emit large amounts of CO_2 into the atmosphere - approximately 350 kg CO_2/MWh for a natural gas combined cycle (NGCC) plant [2]. Coupling power plants with carbon capture and storage (CCS) technologies has been proposed as a promising alternative to decrease CO_2 emissions, whilst continuing to use the vast fossil fuel reserves available worldwide. Several CO_2 capture options have been studied for gas-fired power plants, including pre-combustion, oxy-combustion and post-combustion schemes. This work focuses on post-combustion CO_2 capture in NGCC power plants using amines, which is a capture technology that has already been tested at commercial scale for coal applications [3, 4].

An important challenge for post-combustion CO_2 capture in NGCC power plants is related to the large flow of flue gas with low CO_2 concentration (around 3-4%vol.) generated in the process, which comes as a consequence of the high excess air used in the combustor. This hinders the CO_2 capture process by limiting the driving force in the absorption column, and thus, results in increased capture costs. New process alternatives are currently being investigated to overcome these limitations and optimize CO_2 capture processes in gas-fired power plants. This is the case of exhaust gas recirculation (EGR) [5-9] and selective exhaust gas recirculation (S-EGR) [10] processes. The EGR process is shown in Figure 1. It consists of recirculating a fraction of the flue gas back to the inlet of the compressor (after being cooled down), thereby replacing some of the air fed to the gas turbine. The non-recirculated fraction of the exhaust gas is subsequently treated in a CO_2 capture plant. As a result, the CO_2 concentration of the flue gas sent to the capture system increases and its volumetric flow decreases, which reduces the burden on the CCS plant [5-

9]. However, recirculation of flue gas in the system of Figure 1 also reduces the oxygen available in the combustor, which can affect flame stability, combustion efficiency and emissions. Previous studies [11, 12] have shown that oxygen levels at the inlet of the combustor should not be lower than 16%vol. with current combustor design. This limits the maximum achievable EGR ratios to 40% (i.e, around 6.5%vol. CO_2 in the flue gas [7]).

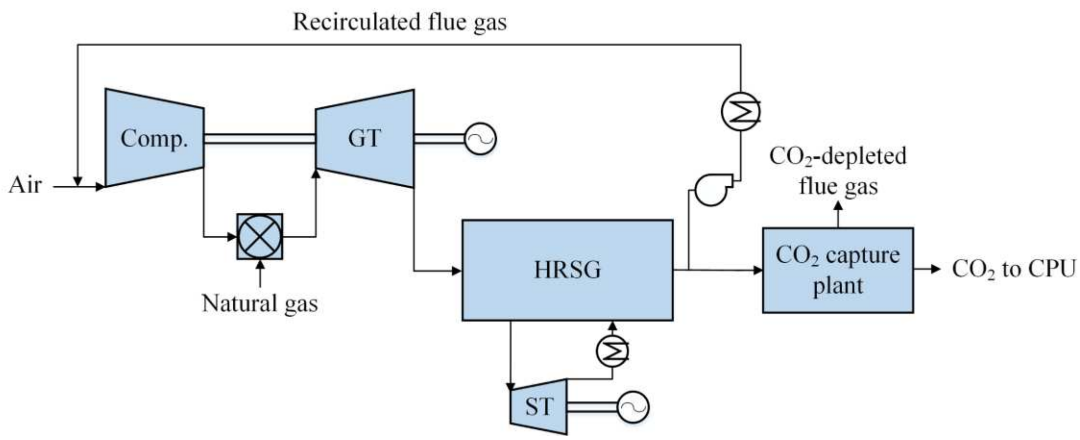

Figure 1. Schematic of the EGR configuration in a NGCC power plant using amines.

To overcome this limitation and achieve a further increase in the CO_2 content of the flue gas, Merkel et al. [10] proposed the application of selective exhaust gas recirculation. The idea behind the S-EGR concept is to selectively recirculate CO_2 upstream of the post-combustion CO_2 capture process, from the gas turbine exhaust back through the inlet of the engine, thus largely increasing the CO_2 concentration of the flue gas sent to the capture plant. For this purpose, S-EGR makes use of a membrane that selectively separates the CO_2 contained within the flue gas. The two S-EGR schemes proposed by Merkel et al. [10] are shown in Figures 2a and 2b, where the membrane and the CO_2 capture plant are arranged in parallel or series, respectively. In both cases, a fraction of the flue gas is passed through a membrane, where mainly CO_2 permeates and mixes with an air sweep stream that flows countercurrently. According to Merkel et al. [10], the CO_2 separation in the membrane is driven by the different CO_2 concentration in the permeate and retentate streams, and no compression or vacuum is required, apart from that needed to overcome the pressure drop in the membrane. The CO_2-enriched air is then introduced to the compressor of the gas turbine, whereas the remaining fraction of the flue gas that has not been recirculated is taken to a CO_2 capture plant. Since the membrane is selective for CO_2, almost no nitrogen or other dilutant species permeate to the oxidant stream, so that higher S-EGR ratios and higher flue gas CO_2 concentrations can be potentially achieved without compromising combustion.

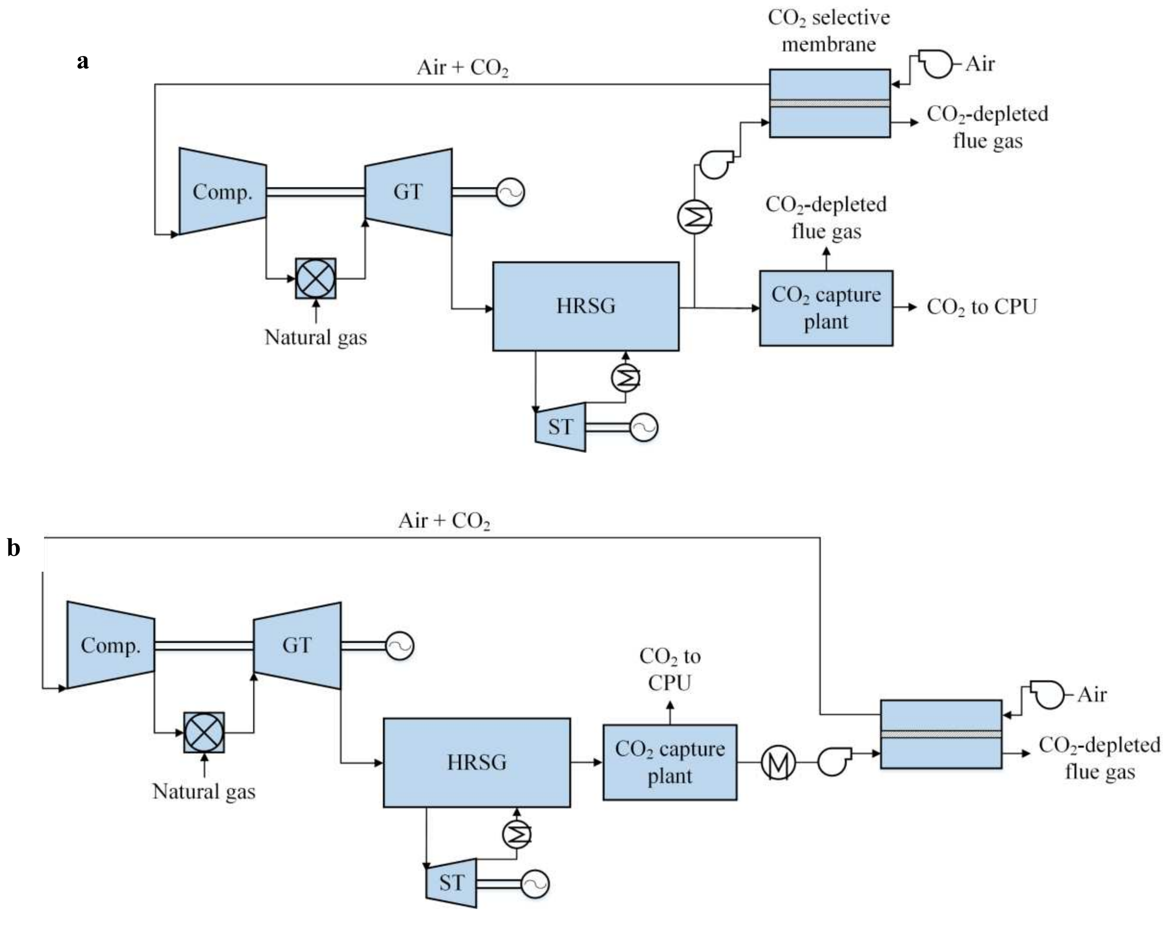

Figure 2. Schematic of the (a) parallel and (b) series configuration in a NGCC power plant.

This study investigates the parallel S-EGR concept of Figure 2a for its application to NGCC power plants. For this purpose, this configuration is simulated and compared to an unabated NGCC system (i.e., no CO_2 capture), a conventional NGCC (without EGR or S-EGR) coupled with an amine capture plant, and an EGR NGCC power plant using amine scrubbing as the downstream capture technology. A preliminary cost analysis has been also carried out to benchmark these options and identify potential opportunities offered by the parallel S-EGR configuration.

METHODOLOGY

The unabated 650 MW_e NGCC power plant simulated in a recent report of the US Department of Energy (DOE/NETL) [13] is taken as reference. It is simulated using gCCS v1.1.0 (PSE), which is a process modelling tool from the gPROMS family that has been specifically designed for power and CO_2 capture systems [14]. The power plant is comprised of two gas turbines (GT) of the GE 7 FA.05 type, two heat recovery steam generators (HRSG) and a steam turbine (ST). The turbine inlet temperature (TIT) is 1360 °C, and the HRSG has three pressure levels with reheat (steam turbine operation: 162/24/3 bar, 564/564/288 °C). The isentropic steam turbine efficiency of the high, intermediate and low pressure sections is 88.0/92.4/93.7 (HP/IP/LP), and the condenser operates at ~0.05 bar. Table 1 shows a summary of the main operating conditions used in this study, which are taken from [13].

Table 1. Main operating conditions of the NGCC power plant.

Compressor pressure ratio	17
Compressor isentropic efficiency (%)	83.7
Pressure drop in the combustor (%)	5
Turbine isentropic efficiency (%)	91.4
Turbine inlet temperature (TIT) (°C)	1360
ST isentropic efficiency (HP/IP/LP) (%)	88.0/92.4/93.7
Air inlet temperature (°C)	15[a]
Fuel inlet temperature (°C)	38
Fuel composition (%vol.)	
CH_4	93.1
C_2H_6	3.2
C_3H_8	0.7
C_4H_{10}	0.4
CO_2	1.0
N_2	1.6
Fuel lower heating value (LHV) (MJ/kg)	47.2
Thermal input power (MW_{LHV})	1103

[a]Air composition (%vol.): 77.32% N_2, 20.74% O_2, 0.92% Ar, 0.03% CO_2, 0.99% H_2O

Modeling Strategy for the NGCC Power Plant with CO_2 Capture

As mentioned above, three different CO_2 capture configurations are simulated in this work, all of them coupled to an amine capture plant (ACP): a conventional NGCC (NGCC+CCS), a NGCC with EGR (EGR), and a NGCC with parallel S-EGR (parallel S-EGR), which is depicted in Figure 3. For this purpose, all operating conditions of the NGCC power plant remained the same as in Table 1. Therefore, the thermal input power used was that of the unabated NGCC case (1103 MW_{LHV}), and the air flow rate was adjusted in the EGR and parallel S-EGR cases to maintain the TIT at 1360°C. The HRSG was designed using pinch points of 10°C for all configurations. Also, the flue gas is assumed to be cooled down to 20°C for the EGR and S-EGR cases, prior to recirculation back to the inlet of the compressor (see Figure 3 for the S-EGR case). The EGR ratio in the NGCC power plant with EGR configuration (defined as the ratio between the volume of recirculated gas and the total volume of flue gas) was set at 39%. This value was calculated to lead to an O_2 concentration at the combustor inlet of 16%vol., thereby avoiding issues related to flame stability and combustion efficiency associated with lower oxygen contents in the combustor [12, 15].

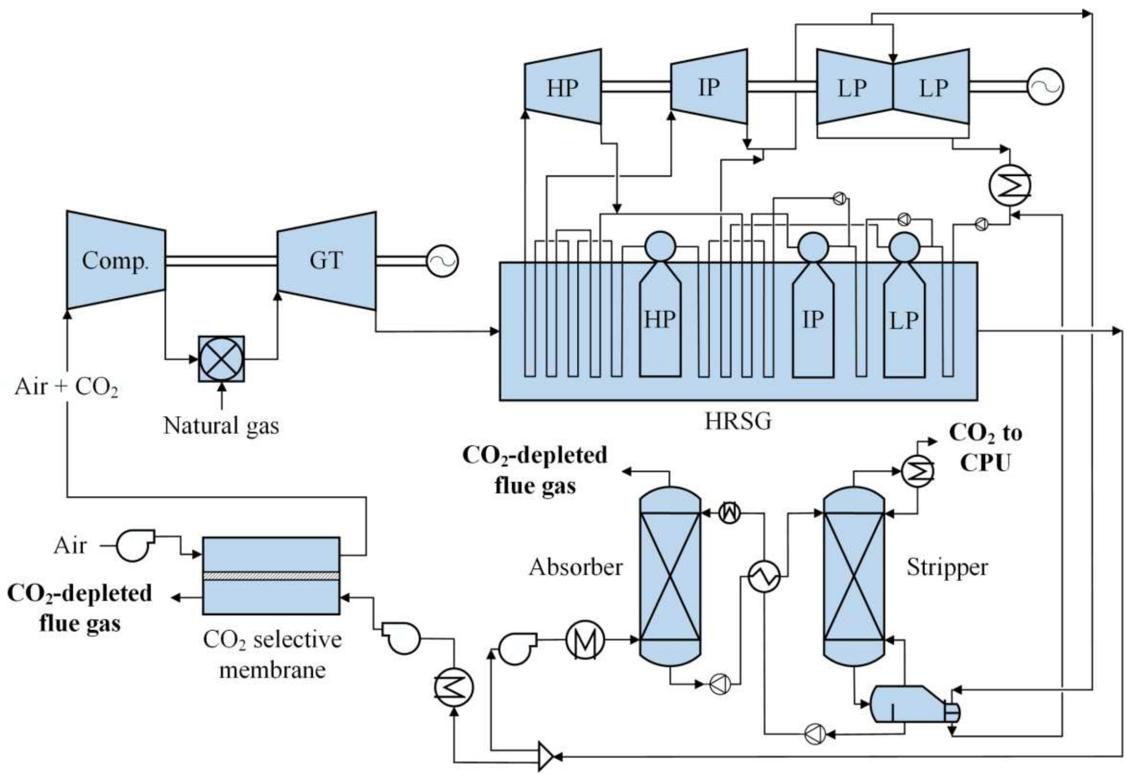

Figure 3. Process scheme of a NGCC power plant using the parallel S-EGR configuration.

In order to calculate the S-EGR ratio to be employed in the NGCC with parallel S-EGR configuration (defined as the ratio between the volume of gas treated in the selective membrane and the total volume of flue gas), an analysis of the evolution of the O_2 concentration at the inlet of the combustor, the CO_2 content of the flue gas and the overall CO_2 capture efficiency was carried out for different S-EGR ratios, as represented in Figure 4. For this purpose, the membrane was assumed to perfectly separate CO_2 from the other gas compounds with an efficiency of 95%, similarly to the value used by Merkel et al. [10]. The amine capture plant was assumed to have a CO_2 capture efficiency of 95%, which can be an optimal value for some applications as discussed in [16].

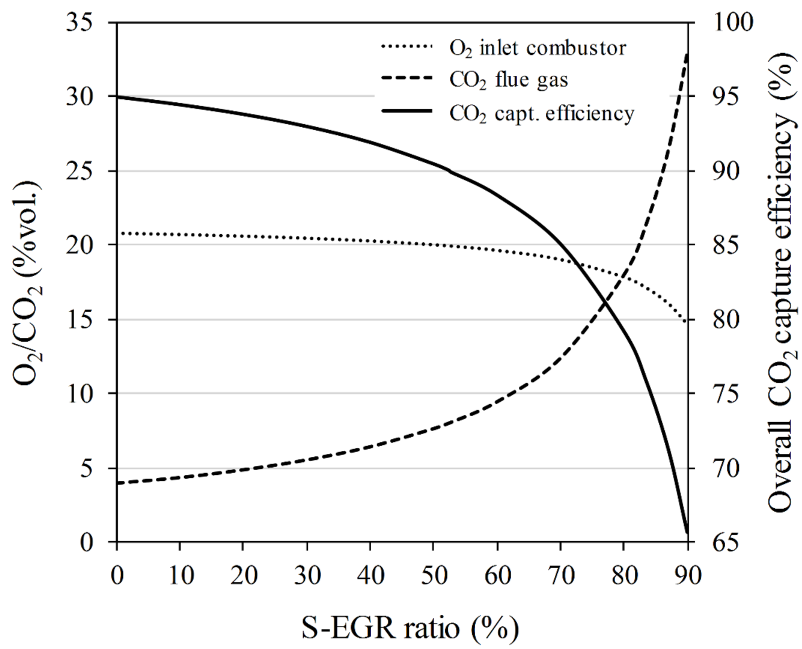

Figure 4. Evolution of the O_2 concentration at the inlet of the combustor, the CO_2 content in the flue gas and the overall CO_2 capture efficiency of the system of Figure 3 with the S-EGR ratio for the parallel S-EGR configuration.

As can be seen in Figure 4, an increase in the S-EGR ratio leads to a progressive increase in the flue gas CO_2 content, whereas the reduction in the O_2 at the inlet of the combustor is almost imperceptible until S-EGR ratios higher than 60% are reached. This is a consequence of the S-EGR configuration, since mainly CO_2 is allowed to pass through the membrane towards the oxidizer (see Figure 3). Another important trend is that followed by the overall CO_2 capture efficiency, which notably decreases with the S-EGR ratio. As the S-EGR ratio increases, a higher flow rate of flue gas with higher CO_2 content is concentrated in the recirculation loop of Figure 3 that passes through the membrane, which is assumed to have a fixed CO_2 separation efficiency in all cases (95%). As a result, the amount of CO_2 that leaves the membrane with the exiting flue gas is higher as the S-EGR ratio increases, and it is emitted into the atmosphere together with the CO_2 depleted gas from the absorber, leading to a reduction in the overall CO_2 capture efficiency of the process. In order to decide the S-EGR ratio to be used in this work, the following two criteria were taken into account: (i) the overall CO_2 capture efficiency of the system is 90%; (ii) the minimum oxygen content at the inlet of the combustor is 16%vol., according to the findings discussed above [12,15]. As can be seen in Figure 4, both conditions are fulfilled when the S-EGR recycle ratio is ~53% and therefore, this was the value selected in this work. It is important to highlight that previous works [10] use a higher S-EGR ratio for the calculations (~77%), whilst achieving 90% overall CO_2 capture efficiency. However, this is because of the very high CO_2 capture efficiencies assumed for the membrane and capture plant individually, which account for 97 and 98%, respectively.

Design of the CO_2 Capture System

The NGCC+CCS, EGR (39% EGR ratio) and parallel S-EGR (53% ratio) configurations studied in this work are designed to achieve an overall CO_2 capture efficiency of 90%. The corresponding amine capture plant was designed and simulated for each case in gCCS v1.1.0, using an aqueous solution of monoethanolamine (MEA) 30%wt. as the solvent. The solvent lean loading was fixed at 0.2 mol CO_2/mol MEA, as it was shown to be optimal in previous studies [17]. The temperature of the flue gas entering the absorber was fixed at 40°C, and the stripper reboiler was taken to operate at 1.75 bar. A summary of all assumptions used in the design of the amine capture plant is included in Table 2.

The amine capture plant was assumed to be composed of two absorbers and one stripper, in order to ensure operational flexibility during part-load scenarios. The design procedure followed for the absorber and stripper columns was similar to that shown in previous works [9, 17]. The column diameter was calculated according to two criteria [9, 17]: (i) a maximum flooding factor of 80%; (ii) a maximum pressure drop across the columns of ~204 Pa/m, which is a recommended value for moderately foaming amine systems [18]. The absorber height was calculated in order to achieve the targeted capture efficiency (90% for the NGCC+CCS and EGR configurations, and 95% for the S-EGR case, as discussed above), selecting an appropriate liquid-gas (L/G) ratio as explained below. The stripper was designed using the reboiler energy consumption as the criterion. For this purpose, the stripper height was progressively increased (using 0.1 m steps). The stripper design height was found when the change in the reboiler duty was considered negligible (lower than 0.05%) with further increase in height. This design procedure was repeated for different L/G ratios in the absorber, which were progressively decreased using 0.01 steps. The final L/G ratio arrived for each configuration when the reduction in the reboiler duty with the associated increase in absorber height was lower than 0.05% with any further decrease in the L/G ratio. It is important to highlight that the estimated L/G ratios are similar to those found in other works for systems with similar flue gas CO_2 concentration (e.g. [9, 17]), thereby validating the procedure described above.

Table 2. Main assumptions for the design of the amine capture plant and membrane system.

Overall CO_2 capture efficiency (%)	90
Pump efficiency (%)	75
Blower adiabatic efficiency (%)	85
Amine Capture Plant	
MEA concentration in the solvent (%wt.)	30
Lean loading (mol CO_2/mol MEA)	0.2
Flue gas temperature at absorber inlet (°C)	40
Flue gas pressure at absorber inlet (bar)	1.13
Lean solvent temperature at absorber inlet (°C)	40
Lean/rich heat exchanger cold outlet approach (°C)	10
Number of absorbers	2
Number of strippers	1
Stripper condenser temperature (°C)	35
Pressure of the stripper reboiler (bar)	1.75
Pressure rich amine pump (bar)	3
Pressure lean amine pump (bar)	3
Column packing	IMTP50
CO_2 Selective membrane	
CO_2 membrane separation efficiency (%)	95
Pressure drop across the membrane (bar)	0.1

Moreover, the CO_2 separation efficiency of the membrane used in the S-EGR case is fixed at 95%, whilst the pressure drop of both the air and flue gas across the selective membrane is assumed to be of 0.1 bar (approx. 10% ΔP) [10].

SIMULATION RESULTS AND DISCUSSION

The simulation results obtained in this study are summarized in Table 3. For the purpose of comparison, this table also contains the results obtained for the 650 MW_e unabated NGCC power plant used as reference (named as NGCC in the table). They match well with those of the DOE/NETL report [13] for the similar case, what validates the NGCC model built in gCCS.

As expected, coupling the power plant with a CO_2 capture amine plant (NGCC+CCS case) leads to a substantial decrease in the power output, thereby reducing the net efficiency (LHV) from 57.5 to 49.0% (see Table 3). It is important to note that the net efficiency of the NGCC+CCS system calculated here is slightly lower than that reported for the same case in the DOE/NETL report (49.0% vs 50.1%) [13]. This is because DOE/NETL uses an advanced solvent in the amine capture plant instead of MEA 30%wt., which reduces the overall energy requirements to regenerate the solvent in the reboiler. As shown in Table 3, the calculated reboiler duty of the NGCC+CCS system is 3.95 MJ/kg CO_2 in this work. This value is in accordance with previous studies on similar amine systems capturing CO_2 from a diluted (3.9%vol. CO_2 in this case) flue gas stream (see for example [8, 17]).

If exhaust gas recirculation is considered, the gas turbine power output reduces slightly, mainly due to the higher temperature of the gas entering the compressor (the recirculated gas is at 20°C) and the slightly decreased flow rate (calculated to maintain TIT at 1360°C). However, more steam is available for generating power in the steam turbine, thereby leading to a higher overall gross power output, as indicated in Table 3. This is because of the higher temperature of the flue gas entering the HRSG, but also due to the lower energy requirements of the capture system, which allows more steam to be available for energy generation in the steam turbine. The reduction in the energy consumption of the amine plant is a direct consequence of the 40% decrease in the flue gas flow rate sent to the amine capture plant, but also of the increase in the CO_2 content (up to 6.5%vol.), which reduces the reboiler duty to 3.74 MJ/kg CO_2. This value matches well with those reported by Li et al. [7], which obtained an equation to express the dependence of the energy demand of the stripper with the CO_2 concentration in a similar amine capture system. In overall terms, the power plant net efficiency of the EGR case increases in 0.8 net percentage points, up to 49.8%, with respect to the NGCC+CCS case. In addition, the total packing volume of the amine capture plant is reduced by 22%, thereby decreasing the total capital expenditure (CAPEX).

Table 3. Simulation results obtained in gCCS for all cases studied.

	NGCC	**NGCC + CCS**	**EGR (39%)**	**Parallel S-EGR (53%)**
GT power output (MW_e)	418.7	418.7	415.0	411.3
ST power output (MW_e)	231.7	174.0	181.1	184.7
Total gross power output (MW_e)	650.4	592.7	596.1	596.0
Power plant auxiliaries (MW_e)	16.3	14.6	14.9	15.0
EGR/S-EGR auxiliaries (MW_e)	-	-	1.1	16.3 (8.6[b])
ACP auxiliaries (MW_e)	-	17.6	11.0	8.9
Compression auxiliaries[a] (MW_e)	-	20.2	20.2	20.2
Total auxiliary loses (MW_e)	16.3	52.4	47.1	60.4 (52.7[b])
Total net power output (MW_e)	634.1	540.2	549.0	535.6 (543.3[b])
Efficiency (%)	57.5	49.0	49.8	48.6 (49.3[b])
CO_2 emissions (kg/MWh_e)	354	42	41	42 (41[b])
Amine capture plant (ACP)				
No. absorbers	-	2	2	2
Absorber height (m)	-	17.1	21.4	22.0
Absorber diameter (m)	-	15.0	11.6	10.4
No. strippers	-	1	1	1
Stripper height (m)	-	26.9	27.3	27.2
Stripper diameter (m)	-	7.6	7.4	7.4
Total packing volume (m^3)	-	7264	5697	4908
Reboiler duty (MJ/kg CO_2)	-	3.95	3.74	3.71
Flue gas flowrate to the ACP (kg/s)	1030	1030	621	488
Flue gas composition to ACP (%vol)				
N_2	74.4	74.4	75.9	71.3
O_2	12.4	12.4	7.7	11.3
Ar	0.9	0.9	0.9	0.8
CO_2	3.9	3.9	6.5	8.0
H_2O	8.4	8.4	9.0	8.6

[a] The energy requirements for the compression and purification unit (CPU) are taken to be 100 kWh_e/t CO_2 [19].
[b] Values assuming a pressure drop across the membrane of 0.05 bar.

The last case in Table 3 is the parallel S-EGR configuration. As can be seen, the power generated in the gas turbine decreases slightly, up to 411.3 MW_e. This is mainly due to the higher temperature of the air+CO_2 mixture entering the compressor, since air is first compressed in a blower (thus increasing its temperature) prior to entering the membrane and is then mixed with the CO_2 gas that permeates from the flue gas, which is at 20°C. Nevertheless, the power output of the steam turbine increases 6 and 2% with respect to the NGCC+CCS and EGR configurations, respectively. As in the ERG case, this is due to the higher temperature of the flue gas entering the HRSG and the lower steam requirements of the amine capture plant, which treats a flue gas flow rate 53% and 21% lower than in the NGCC+CCS and EGR cases, respectively, with 8%vol. CO_2. As a result, the total gross power output is almost identical to that of the EGR scheme, and higher than that of NGCC+CCS. According to the discussion above, the total packing volume calculated for the amine capture plant reduces by 32% and 14% of that in NGCC+CCS and EGR cases, respectively, and the reboiler duty is 3.71 MJ/kg CO_2, following a similar trend as that in [7]. As a result, the amine capture plant auxiliary consumption is reduced (see Table 3). However, a substantial increase in the S-EGR auxiliaries is calculated, due to the need of an air and a flue gas blower to overcome the pressure drop in the membrane. As a result, the calculated net NGCC efficiency for the parallel S-EGR configuration decreases to 48.6%. It should be noted that any reduction in the pressure drop through the membrane will highly benefit the parallel S-EGR configuration studied here. In order to study these variations, Table 3 also contains the values obtained for this case when the pressure drop across the membrane is assumed to be ~5% (i.e., 0.05 bar). As can be seen (values in brackets in the table), the total net power output in this case is closer to that in the EGR case, and higher than in the reference NGCC+CCS, thereby leading to a net electrical efficiency of 49.3%.

Moreover, CO_2 emissions from the NGCC power plant reduce significantly for all the CCS configurations studied (from ~350 to ~40 kg CO_2/MWh_e), as indicated in Table 3. The results obtained indicate that EGR has the lowest efficiency penalty of the configurations studied in this work. However, these calculations also show that the parallel S-EGR configuration may also be competitive, especially if the power consumed by the auxiliaries of the S-EGR system could be reduced. Additionally, further benefits can be derived from the S-EGR system with regards to the amine capture plant. These include reductions in size of the main reactors of the capture system (absorber, stripper, direct contact cooler) due to the lower flue gas flow rate compared to EGR, and of the equipment (e.g. blower). Also, more compact designs could arise as a result of the increased CO_2 content in the flue gas. In order to quantify these effects, a high-level cost analysis has been carried out, as will be described below.

PRELIMINARY COST ANALYSIS

An initial study of the preliminary costs associated with the NGCC configurations outlined above has been performed. This study uses the capital cost scaling methodology outlined by NETL [20] to perform a high level cost evaluation and DOE/NETL 'Case 1B 7FA.05' as the reference plant [13], which is a NGCC plant with two gas turbines, heat recovery steam generators (HRSG) and one steam turbine with an amine CO_2 capture plant. The cost of electricity (COE) and cost of CO_2 avoided (COA) are the two financial parameters considered in this analysis, as defined by IPCC [21]. The COE (\$/MWh) is determined by:

$$COE = [(TOC)(CCF)+(FOM)]/[(CF)(8760)(MW)] + VOM + (HR)(FC) \tag{1}$$

where CCF is the capital charge factor, CF is the capacity factor and MW is the net plant power (MW_e). The total overnight costs (TOC), fixed operating and maintenance (FOM) costs and variable operating and maintenance (VOM) costs are expressed in US\$, \$/yr and \$/MWh, respectively. The net heat rate (HR) and fuel cost (FC) are expressed in MJ/MWh and \$/MJ, respectively.

The COA (\$/ton CO_2 avoided) can be expressed by:

$$COA = (COE_{CCS} - COE_{REF}) / (EMS_{REF} - EMS_{CCS}) \tag{2}$$

where COE_{CCS} and COE_{REF} are the cost of electricity with and without CO_2 capture, respectively. In the above equation, EMS_{CCS} and EMS_{REF} represent the CO_2 emission rate with and without CO_2 capture. The COE_{REF} and EMS_{REF} are illustrated in Table 4.

Table 4. Economic assumptions.

Description	Value	Unit
Capacity factor (CF) [(a)]	85.0	%
Capital charge factor (CCF) with CCS [(a)]	0.11	Fraction
COE_{REF} [(a)]	57.1	$/MWh_e$
EMS_{REF} [(a)]	354	kg CO_2/MWh_e
CO_2 transport and storage over 100 km [(a)]	10	$/ton
Natural gas price [(a)]	5.8	$/GJ
Plant operating period [(a)]	30	Years
Membrane installed skid cost [(b)]	50	$/m^2$
Membrane module lifetime [(c)]	5	yr
Total as spent cost (TASC) multiplier [(a)]	1.078	Fraction
Cost year [(a)]	2011	-
Labor rate [(a)]	51.6	$/h
Labor per shift [(a)]	6.3	-
Shifts per day [(a)]	3	-

[(a)] Economic assumptions from reference [13].
[(b)] Membrane skid cost from references [10, 22].
[(c)] Membrane lifetime from reference [22].

The main economic assumptions used in this study to calculate COE and COA for the different configurations are summarized in Table 4.

Capital costs

The main contributor to the total overnight cost (TOC) included in equation 1 is the total plant cost (TPC). This cost has been calculated in this work using the values of the DOE/NETL report for the NGCC with CO_2 capture case as the reference (‘Case 1B 7FA.05’) [13]. For this purpose, the cost of each major process area outlined in Figure 3 is calculated following the methodology in [20]. To consider the differences in scale arising in each configuration, the following expression has been used [20]:

$$SC = RC. (SP/RP)^{Exp} \tag{3}$$

where SC is the scaled cost, RC is the reference cost, SP and RP stand for the scaling and the reference parameter, respectively, and Exp is the scaling exponent.

The scaling parameters and exponents used to calculate the scaled costs for each process area of each configuration (employing equation 3) are presented in Table 5. As can be seen in the table, the cost of the CO_2 removal system for the different configurations is scaled using the flue gas and solvent flow rates. This has been carried out according to the procedure in [23], which indicates that the flue gas entering the absorber should be used to scale the costs of the direct contact cooler, blower and absorber units (named here as the absorption section) of the amine capture system, whereas the solvent flow rate is the scaling parameter for all solvent-related units (heat exchangers, circulation pumps, stripper, reboiler – i.e., the solvent-related section). For this purpose, the contribution of the absorption and solvent sections to the total reference costs of the CO_2 removal system was taken to be 65 and 35%, respectively, as estimated from the IECM v9.2.1 software [24].

Table 5. Scaling parameters.

Process area	Parameter description	Scaling exponent [(a)]
Feed water system	HP feed water flow rate	0.72
Natural gas pipeline	Fuel gas flow rate	0.07
Natural gas miscellaneous	Fuel gas flow rate	0.76
EGR system	EGR flow rate [(b)]	0.70
CO_2 removal system – absorption section	Flow rate to absorber inlet	0.61
CO_2 removal system – solvent-related section	Rich solvent flow rate	0.61
CO_2 comp. and drying	CO_2 flowrate	0.77
HRSG system	HRSG duty	0.70
Steam turbine system	Steam turbine power	0.80
Cooling water system	Cooling tower duty	0.71

[(a)] Values from scaling methodology [20].
[(b)] The EGR flow rate value for the reference plant is from the DOE/NETL report with the EGR system ('Case 1C 7FA.05') [13].

It is important to note that in order to calculate the costs of the membrane, the area required needs to be estimated. For this purpose, a similar procedure as in [25] was used, dividing the membrane in 10 steps of equal area and assuming a CO_2 permeance of 2200 gpu, as well as perfect separation of CO_2. The parallel membrane system cost is then determined as a function of the membrane area (A_M) and the membrane installed skid cost (C_{skid}), which is assumed to be of 50 $/m^2, as in [10, 22]:

$$\text{Membrane capital cost} = A_M .\, C_{skid} \tag{4}$$

The permeance value considered in this study is that indicated in [10] for the state-of-the-art Polaris membrane, which has a selectivity of ~50-60 [10, 26]. Due to the characteristics of the S-EGR system (use of an air sweep stream, no operation under pressure/vacuum [10]), the concentration gradient of non-CO_2

gases (N_2 and O_2 mainly) between the permeate and retentate streams in the membrane is limited, so high selectivity is not expected to be as crucial as in other membrane applications.

In addition to the total plant cost, Table 6 outlines the other economic parameters which contribute to the total overnight cost. As indicated in Table 6, these have been calculated assuming the values from DoE [13] for the NGCC plant coupled with an amine capture system.

Table 6. Other economic parameters for TOC.

Item	Parameter
Preproduction costs [(a)]	6 months operating labor 1 month maintenance materials cost 1 month non-fuel consumables 25% of 1 month fuel cost 2% TPC
Inventory capital [(a)]	2 months non-fuel consumables Spare parts (0.5% TPC)
Other [(a)]	Initial cost for chemicals (0.002% TPC) [(b)] Land Other owners costs (15%TPC) Financing costs (2.7% TPC)

[(a)] Economic assumptions from reference [13].
[(b)] Economic assumption calculated by the authors from [13].

Operating and Maintenance (O&M) Costs

The annual O&M costs include fixed and variable costs, which are determined from the assumptions presented in Table 7.

Table 7. Operating and maintenance economic parameters.

Item	Parameter
Fixed O&M costs [(a)]	Annual operating labor (AOL) (determined from Table 4) Annual maintenance labor (AMC) (40% of total maintenance costs)[(b)] Administrative and support labor (25% of AOL+AMC) Tax and insurance (2% TPC)
Variable O&M costs	Maintenance material cost (1.1% TPC) [(c)] Non-fuel consumables cost (0.6% TPC) [(c)] Membrane replacement cost (determined from assumptions in Table 4 - 10 $\$/m^2yr$)

[(a)] Economic assumptions from reference [13].
[(b)] Economic assumption from reference [27].
[(c)] Economic assumption calculated by the authors from [13].

In addition to other common O&M costs for all the configurations, the S-EGR system studied in this work requires to periodically replace the membranes. A membrane module lifetime of five years has been assumed in this study (see Table 4), as in Zhai and Rubin [22], leading to a membrane replacement cost of 10 $\$/m^2yr$ as indicated in Table 7.

Economic Analysis and Results

The results of the economic analysis explained above are illustrated by Table 8, which represents the cost for each process area from the reference case (Ref.) [13] and for the different NGCC plant configurations investigated in this study. It is important to highlight that according to [13, 20, 28], the scaling methodology used to calculate costs is only suitable for high level assessment and the economic results presented can reflect up to ±30% uncertainty. It is also important to mention that the economic methodology [20] used in this study has been validated against the different NGCC configurations outlined in the DOE/NETL report [13]. Therefore, the calculation procedure used in this study is considered robust and repeatable. An example of this validation can be seen in Table 8, which shows that the total plant cost for the NGCC+CCS case is very similar to that of the reference plant used in the economic analysis (same configuration as in NGCC+CCS), with only 1% error in the calculated total overnight costs. Also, the differences found in the COE and COA between these configurations are mainly due to the slightly lower power output calculated for the NGCC+CCS configuration (540 MW_e (see Table 3) vs 553 MW_e [13]). This is due to the higher energy penalty of the amine capture plant in the NGCC+CCS system as a result of the use of MEA 30%wt. as solvent instead of an advanced blend [13], as discussed above.

Table 8. Preliminary cost analysis results.

		Ref. plant [a]	NGCC + CCS	EGR (39%)	Parall. S-EGR (53%)
	Process area	(M$)			
1	Feed water system and natural gas pipeline	55.8	56.0	56.7	57.3
2	EGR system [b]	-	-	21.5	26.5
3	CO_2 removal system[c]	313.1	311.2	260.0	242.4
4	Gas turbine system	134.0	134.0	134.0	134.0
5	HRSG system	55.5	54.9	55.5	56.1
6	Steam turbine system	66.9	63.6	65.6	66.7
7	Cooling water system	26.2	20.9	21.9	22.2
8	Parallel membrane system	-	-	-	49.9
9	Accessory electric plant, instrumentation & control, improvements to site, buildings & structures (14% TPC)	107.1	105.3	101.2	107.7
	Total plant cost (TPC)	**758.7**	**746.0**	**716.5**	**762.9**
10	Preproduction costs	26.4	25.9	25.3	26.2
11	Inventory capital and initial chemical cost	5.8	5.9	5.6	6.0
12	Land costs	0.3	0.3	0.3	0.3
13	Other owners cost (15% TPC)	113.8	111.9	107.5	114.4
14	Financing costs (2.7% TPC)	20.5	20.1	19.3	20.6
	Total overnight cost (TOC)	**925.5**	**910.1**	**874.5**	**930.4**
	Total as spent cost (TASC)	**997.7**	**981.0**	**942.7**	**1003.0**
15	Total FOM	25.7	22.6	21.8	27.7
16	Total VOM	12.9	12.7	12.2	22.4
17	Total fuel cost	190.5	190.5	190.5	190.5
	Total O&M cost	**229.1**	**225.8**	**224.5**	**240.5**
	CO_2 transport and storage cost ($/$MWh_e$)	**3.7**	**3.7**	**3.7**	**3.8**
	COE ($/$MWh_e$)	**84.3**	**85.0**	**82.4**	**90.0**
	COA ($/ton CO_2 avoided)	**86.6**	**89.0**	**80.4**	**105.1**

[a] Values from reference [13].
[b] This includes the EGR blowers, cooler and cooler pumps [13].
[c] This includes the cost of the amine capture plant and of the CO_2 compression and drying system [13].

The results outlined in Table 8 show that the total plant cost varies for the EGR and S-EGR cases by -4% and +2%, respectively, compared to the NGCC+CCS plant. The primary differences in total plant cost are associated with the membrane costs in the S-EGR configuration, and with the lower capital costs for the CO_2 removal system in both the EGR and S-EGR cases. In fact, the calculated capital costs (including the cost of the amine capture plant and the CO_2 compression and drying system) reduce by 16% (EGR case) and 22% (parallel S-EGR case), respectively, compared to the NGCC+CCS plant costs. The main reduction in the CO_2 capture costs is related to the absorption section (direct contact cooler, blower, absorber), whereas the cost of the solvent-related section (heat exchangers, circulation pumps, stripper, reboiler) remains similar for all configurations, as expected from the minor changes in the stripper size (and associated solvent flows and equipment) shown in Table 3. In both EGR and parallel S-EGR a proportion of the flue gases is recirculated which reduces the flowrate to the inlet of the absorber (see Table 3), decreasing the overall unit cost, as the size reduces. The flowrate to the inlet of the absorber is the parameter used to calculate the costs of the absorption section in the CO_2 capture system, as illustrated in Table 5. In addition, the CO_2 concentration in the flue gas stream increases as the recirculation ratio increases. This allows the mass transfer of separating CO_2 from the other flue gases in the absorption column to be optimized. In the S-EGR case, the parallel membrane system separates CO_2 from the other flue gases. This allows higher recirculation ratios to be achieved compared to the EGR case, further reducing the cost of CO_2 removal.

The results from this study show that, although the S-EGR configuration offers benefits regarding the CO_2 removal system, the cost of the parallel membrane system increases the total plant cost to around 763 M$, compared to 716 M$ for the EGR case, representing a 6% increase. The total operating and maintenance costs calculated for the parallel S-EGR configuration are also 7% higher, which can be mainly attributed to the membrane replacement cost and the higher total plant costs (see the calculation procedure in Table 7).

In comparison to the NGCC+CCS plant, the COE and COA for the EGR configuration reduce by 3% and 10%, respectively, to 82.4$/MWh and 80.4 $/ton CO_2 avoided, respectively. However, an increase of 6% and 18% for the COE and COA for S-EGR is observed. These variances are attributed to parallel membrane system capital, fixed and operating costs, and the reduction of net power output (see results in Table 3), which is mainly due to the higher calculated S-EGR auxiliary consumption.

Sensitivity Analysis

The initial results indicate that parallel S-EGR may be more costly compared to the other cases investigated, however, there are other factors that should be considered to determine if this technology can become more competitive. Membrane systems for post-combustion CO_2 capture are still under development, and their performance and cost after incorporating new advancements is still unclear. The membrane module reference cost 50 $\$/m^2$ has been widely reported in the literature e.g. [10, 22, 25, 29]. As parallel membrane systems are improved, variation to membrane costs and technical improvements are anticipated. This will be of high benefit for the parallel S-EGR configuration studied in this work amongst the NGCC+CCS and EGR schemes. In addition, pressure drop across the membrane plays an important role (assumed to be 0.1 bar in this work, i.e., approx. 10%) in terms of the energy auxiliary consumption in the parallel S-EGR case, as seen in Table 3, which reduces the total net power output, thereby increasing COE and COA for this configuration. It is also important to investigate the potential of the S-EGR configuration to further reduce the cost of the amine capture plant. As discussed above, this is not only due to the reduction in the flue gas flowrate (which is the scaling parameter taken into account in the economic analysis for the absorption section – see Table 5), but also to the increase in the CO_2 concentration of the exhaust gas fed to the absorber, which can lead to further advantages for the capture plant in terms of reduction of the running costs and/or compact equipment design. However, the effect of the enhanced CO_2 concentration is not easy to quantify in an economic analysis.

Table 9. Sensitivity analysis scenarios.

	Scenario						
	a	**b**	**c**	**d**	**e**	**f**	**g**
Membrane[(a)]	1	0.5	0.25	1	1	0.5	0.25
Capture plant S-EGR[(a)]	1	1	1	0.8	0.8	0.8	0.8
ΔP membrane (bar)	0.1	0.1	0.1	0.1	0.05	0.05	0.05

[(a)] Fraction of the cost considered in each scenario with respect to that in Table 8.

To determine what impact any improvements to the techno-economic aspects of parallel membrane systems has on the COE and COA, a sensitivity analysis has been performed. In total, seven scenarios (a-g) are considered, whereby, variations to the cost of the membrane and CO_2 removal system and to the pressure drop across the membrane are implemented. The different scenarios are presented in Table 9, and consider membrane cost 0.5 and 0.25 times that of the reference cost and 0.8 times that of the CO_2 removal system cost in Table 8. Furthermore, the pressure change across the membrane varies from 0.1 to 0.05 bar (from approx. 10 to 5% ΔP). The results of the sensitivity analysis are presented in Figures 5 and 6.

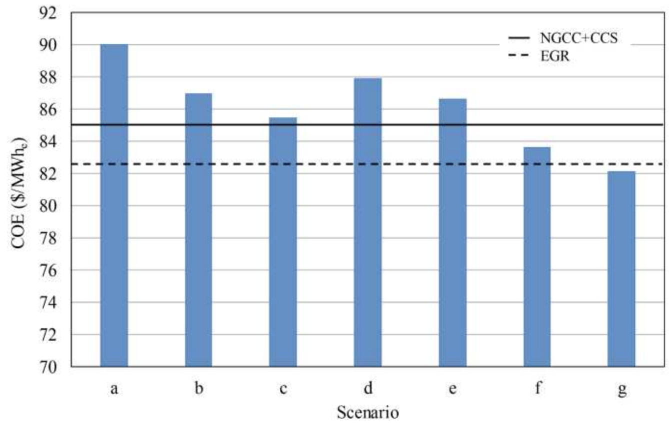

Figure 5. Sensitivity analysis on COE for the S-EGR plant under different scenarios which alter membrane cost, CO_2 capture plant cost and pressure difference across the membrane (black bold and dashed lines represent the costs of the NGCC+CCS and EGR systems, respectively).

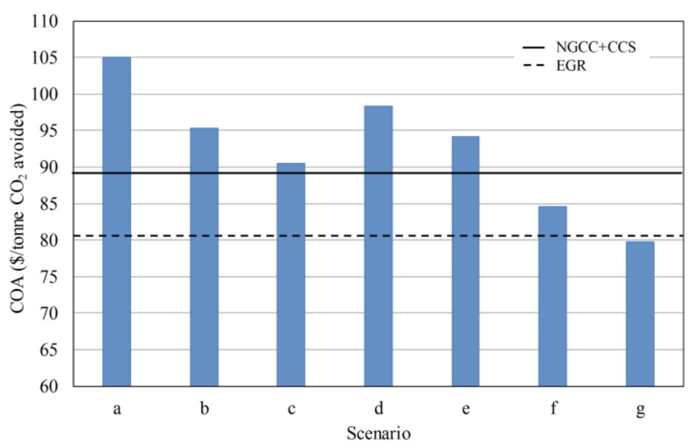

Figure 6. Sensitivity analysis on COA for the S-EGR plant under different scenarios which alter membrane cost, CO_2 capture plant cost and pressure difference across the membrane (black bold and dashed lines represent the costs of the NGCC+CCS and EGR systems, respectively).

As expected, a combined reduction in the pressure difference, membrane cost and CO_2 removal system cost has the greatest influence on the COE and COA compared to the NGCC+CCS case, as shown in Figures 5 and 6. Reducing the membrane cost by 50%, CO_2 removal system by 20% and halving the pressure difference across the membrane (scenario f) decreases the COE and COA by 2% and 5%, respectively, to 83.6 \$/MWh and 84.5 \$/ton CO_2 avoided, compared to the NGCC+CCS case. It is also clear that any additional cost reduction in the parameters mentioned above, will be of benefit for the S-EGR configuration, like in scenario g. Furthermore, compared to the EGR case, these changes make the parallel S-EGR systems more competitive.

CONCLUSIONS

An analysis of the parallel S-EGR configuration for CO_2 capture in NGCC power plants has been performed, assuming 53% S-EGR ratio. The results obtained indicate that the net efficiency achieved by the parallel S-EGR system is between 48.6 and 49.3%, depending on the pressure drop across the membrane (~ 10 and 5%, respectively), compared to 49.0% and 49.8% for the NGCC+CCS and EGR (39% ratio) configurations, respectively. This information together with available costing data has been used to perform a high-level economic analysis, in order to study the effect of key variables (membrane cost, S-EGR auxiliary power consumption (i.e., pressure drop of the gases across the membrane) and potential reductions in the cost of the attached amine capture plant) on the cost of the parallel S-EGR system. These results show that the cost of electricity (COE) of the parallel S-EGR configuration is of 90.0 \$/MWh_e for the reference case, although it can vary in the range of 82.1-87.9 \$/MWh_e with combinations of the parameters indicated above. A similar trend is obtained for the cost of CO_2 avoided, which shows large variations in between 79.7 and 105.1 \$/ton CO_2 avoided. The results obtained in this study indicate that the parallel S-EGR system may be competitive with the NGCC+CCS configuration in some of the scenarios studied, in which the cost of membrane systems and auxiliary losses (i.e., due to pressure drop) decrease with new membrane developments. Additional cost advantages will arise from reductions in the CO_2 capture system as a result of the S-EGR conditions. However, further benefits with respect to the EGR configuration will depend on future advancements and cost reductions achieved on membrane-based systems.

NOMENCLATURE

ACP	Amine capture plant
AMC	Annual maintenance labor (M$)
AOL	Annual operating labor (M$)
CCF	Capital charge factor
CF	Capacity factor
COA	Cost of CO_2 avoided ($/ton CO_2 avoided)
COE	Cost of electricity ($/MWh$_e$)
CPU	Compression and purification unit
EGR	Exhaust gas recirculation
EMS	CO_2 emission rate kg CO_2/MWh$_e$)
Exp	Scaling exponent for cost estimation
FC	Fuel cost ($/MJ)
FOM	Fixed operating and maintenance costs (M$)
GT	Gas turbine
HP	High pressure
HR	Heat rate (MJ/MWh)
IP	Intermediate pressure
LP	Low pressure
MEA	Monoethanolamine
MW	Net plant power (MW$_e$)
RP	Reference parameter to scale costs
S-EGR	Selective exhaust gas recirculation
SC	Scaled cost (M$)
SP	Scaling parameter for cost estimation
ST	Steam turbine
TASC	Total as spent cost (M$)
TIT	Turbine inlet temperature (°C)
TOC	Total overnight cost (M$)
TPC	Total plant cost (M$)
VOM	Variable operating and maintenance costs (M$)

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