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Techno-economic assessment of a commercial natural gas combined cycle with a chemical absorption plant using lean vapor compression modification

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Abstract

The primary challenge in integrating post-combustion CO_2 capture (PCC) with natural gas combined cycle (NGCC) is the significant energy consumption and capital costs. The novelty of this paper lies in proposing for the first time an advanced novel configuration that combines lean vapor compression (LVC) for the PCC plant with the NGCC plant incorporating exhaust gas recirculation (EGR) and selective exhaust gas recirculation (SEGR). The simulation results illustrated that implementing 33% EGR can increase the CO_2 level in exhaust gas from a baseline of 4.2 mol% to 6.3 mol%. In comparison, 53% SEGR increased the CO_2 concentration in the flue gas to 8.8 mol%. Among the different configurations examined, SEGR + LVC achieved the highest energy saving for reboiler duty, which was 14 % compared to the baseline. In contrast, the EGR + LVC recorded the highest enhancement in thermal efficiency by 0.7 % points compared to the reference case. The LVC alone resulted in approximately 0.4 % points improvement in thermal efficiency for all configurations evaluated when the gas turbine loads were reduced from 100% to 60%. This indicates that LVC is effective under partial loads. Finally, SEGR + LVC results in the greatest cost reduction for the PCC plant equipment, lowering the cost by 26% compared to the baseline. However, the SEGR has the highest total plant cost and total overnight cost due to additional costs for the CO_2 membrane separation system.

Keywords: natural gas combined cycle; exhaust gas recirculation; selective exhaust gas recirculation; lean vapor compression; post-combustion CO_2 capture

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Abbreviations

CCGT	Combined cycle gas turbine
CEPCI	Chemical Engineering Plant Cost Index
DOE	U.S. Department of Energy
DCC	Direct contact cooler
EGR	Exhaust gas recirculation
ENRTL	Electrolyte Non-Random Two Liquid
GT	Gas turbine
HRSG	Heat recovery steam generator
HP	High pressure
IP	Intermediate pressure
LP	Low pressure
LVC	Lean vapor compression
LHV	Lower heating value
MEA	Monoethanolamine
NETL	U.S. National Energy Technology Laboratory
NGCC	Natural gas combined cycle
PCC	Post-combustion carbon capture
RK	Redlich-Kwong
SEGR	Selective exhaust gas recirculation
SRD	Specific reboiler duty
ST	Steam turbine
TPC	Total plant cost
TOC	Total overnight cost

1. Introduction

Post-combustion carbon capture (PCC) is a promising way to decarbonize the natural gas combined cycle (NGCC). However, capturing CO_2 from the NGCC plant required high energy consumption due to the low CO_2 concentration and high mass flow rate of the flue gas [1]. Research indicates that integrating the PCC into the NGCC plant results in a significant reduction in thermal efficiency from 55.8% to 46.8% [2]. This efficiency loss results from the fact that the PCC plant requires auxiliary loads such as pumps, a flue gas air blower, a CO_2 compression unit, and solvent regeneration duty.

In the literature, extensive efforts have been made to mitigate the energy penalty associated with integrating the PCC plant with the NGGC plant through various process modifications. For example, a study by Sammak et al. [3] introduces the concept of exhaust gas recirculation (EGR) for an NGCC plant. This technique involves recycling a fraction of the flue gas back to the gas turbine cycle. The results [3] showed that applying the 30% EGR enhanced thermal efficiency by up to 0.6% and resulted in a 7% reduction in carbon capture plant cost. A recent report published by the United States Department of Energy [4] demonstrated that implementing 30% EGR increases the CO_2 concentration in flue gas from 4.1% to 5.8%, while simultaneously reducing the feed flow rate to the absorber by 30 %. However, the recycle ratio of EGR is limited to maintain the inlet oxygen level in the combustor at 16% mol [5] to avoid problems such as flame instability and incomplete combustion.

To overcome the oxygen limitation in the combustor associated with EGR, some authors [6,7] suggested a novel approach called selective exhaust gas recirculation (SEGR) for an NGCC plant. In this approach, CO_2 is selectively recycled from the exhaust gas back into the gas turbine cycle. The findings presented by Asadi et al. [6] demonstrated that implementing 77% SEGR was able to increase the CO_2 level in flue gas from around 4% to 18%. This results in a reduction in reboiler duty and total packing volume for the PCC plant by up to 7% and 44%, respectively.

A study by Qureshia et al. [7] proposed various configurations for the SEGR, including parallel, series, and hybrid. These configurations were investigated at partial loads of NGCC with the PCC plant. However, the main downside of implementing SEGR is the large membrane area required to separate CO_2 from other gases, which is estimated to be around 3 million m² when using 77% SEGR [6].

Several investigators [6, 8-10] have investigated an innovative approach that utilizes solar energy to supply the heat required for solvent regeneration in the PCC plant rather than extracting steam from the NGCC plant. This approach enhances the thermal efficiency of the NGCC integrated with the PCC plant from 41.3% to 46.8% [10]. However, the main obstacle to this method is the significant capital costs. The estimated cost is approximately $194.8 M [6] for the solar equipment, which includes solar collectors and thermal storage.

Various researchers [11-13] proposed modifying the PCC plant with an advanced flash stripper (AFS) to recover wasted energy from the top of the stripper before it condensed. However, implementing this modification introduces considerable complexity as it requires adding two heat exchangers and cold and warm bypasses to the conventional CO_2 capture plant.

Numerous studies [14-16] suggested modifying the PCC plant by using absorber intercooling. This technique can help to reduce the temperature profile inside the absorber, resulting in a better absorption rate and lower solvent regeneration duty. The main limitation of this modification is that it offers minor energy savings for the PCC plant, amounting to up to 4% [15] when monoethanolamine (MEA) is used as solvent.

Multiple studies [17-18] proposed enhancing the PCC plant by rich solvent preheating (RSP). In this approach, an additional heater was used to preheat the rich solvent further after it leaves the rich-lean heat exchanger and before it feeds the top of the stripper. However, this modification does not provide any energy savings because there is no internal heat available to supply the additional heater duty.

Many researchers [19-21] have investigated introducing lean vapor compression (LVC) to enhance the performance of a PCC plant integrated with a coal power plant. However, in the context of the NGCC plant, there are limited studies that discuss LVC for the PCC plant utilizing MEA. For example, Arshad et al. [15] reported that LVC can reduce the reboiler duty and the cost by up to 19% and 5%, respectively. Baudoux et al. [22] demonstrated that LVC reduces the reboiler duty by up to 15%. Hosseinifard et al. [23] investigated a combination of various modifications for the PCC plant, including lean vapor compression, solvent split flow, rich solvent recycles, and rich vapor compression. However, all these studies [15, 22-23] were carried out when NGCC operates with a low CO_2 concentration of around 4% in the flue gas. Additionally, these studies [15, 22-23] lack details as the LVC was investigated at a single operational condition. Furthermore, most of these studies lack economic analysis [22-23].

From the above discussion, it's clear that extensive work was conducted using various advanced process modifications to minimize the energy penalty associated with the NGCC coupled with the PCC plant. The above analysis illustrated that EGR, SEGR, and LVC are promising modifications. To the best of the authors' knowledge, no study has investigated the combination of LVC with EGR and SEGR. Therefore, the novelty and goals of this study are summarized in the following points:

I. Compared the effect of different configurations, including baseline, EGR, SEGR, baseline + LVC, EGR+LVC, SEGR+LVC, on carbon capture plant performance.

II. Investigated the effect of partial load conditions of the NGCC on the reboiler duty, thermal efficiency, and power output for the proposed configurations.

III. Perform economic analysis for the suggested configurations.

2. Methodology

2.1 Process description of the NGCC plant with the PCC plant

The 751 MW NGCC plant was simulated using Aspen Plus V14.1, following the design parameters provided by the U.S. Department of Energy (DOE/NETL May 2023) report, case CC2A-F [24]. It should be noted that this report does not include the carbon capture plant. The main reason for selecting DOE/NETL May 2023 is that it provides data on partial loads of the NGCC. This enables the accurate validation of the power plant under partial load conditions (see section 3).

For the carbon capture plant and EGR, case B31B.90 from the DOE/NETL October 2023 report [4] was used. The selected case has a capacity of 740 MW, which is approximately the same as the above power plant chosen (751 MW).

Fig 1 illustrates the process flowsheet of the NGCC plant with the PCC plant. The NGCC plant operates by compressing air and mixing it with fuel in the combustor. This combustion process produces a high-temperature mixture that drives the gas turbine to generate electricity. The remaining heat from the gas turbine's exhaust gas is recovered by the heat recovery steam generator (HRSG), which creates steam at three different pressure levels. This steam drives an additional turbine to generate more electricity, improving the power plant's overall performance.

The exhaust gas leaves the NGCC plant at atmospheric pressure, which means an air blower is required to overcome the pressure drop and push the flue gas into the PCC plant. Following this, the flue gas must be cooled using a direct contact cooler (DCC) to reduce the operating temperature of the absorber column, thereby boosting the absorption rate. After that, the flue gas enters the absorber tower, and therefore, the chemical reaction between the solvent and CO_2 occurs. The CO_2 captured leaves the absorber from the bottom as a rich solvent, while the clean gas is released from the top of the absorber.

The rich solvent is then pumped to the rich-lean heat exchanger to recover heat from the hot lean solvent that leaves the bottom of the stripper. Following this, the heated rich solvent is introduced at the top of the stripper. At the bottom of the stripper, a reboiler is located, which increases the temperature inside the stripper column to facilitate the separation process of CO_2 from the solvent. Finally, the remaining solvent is pumped back to the absorber for reuse, whereas the captured CO_2 is directed to the CO_2 compression unit to prepare it for transportation and storage.

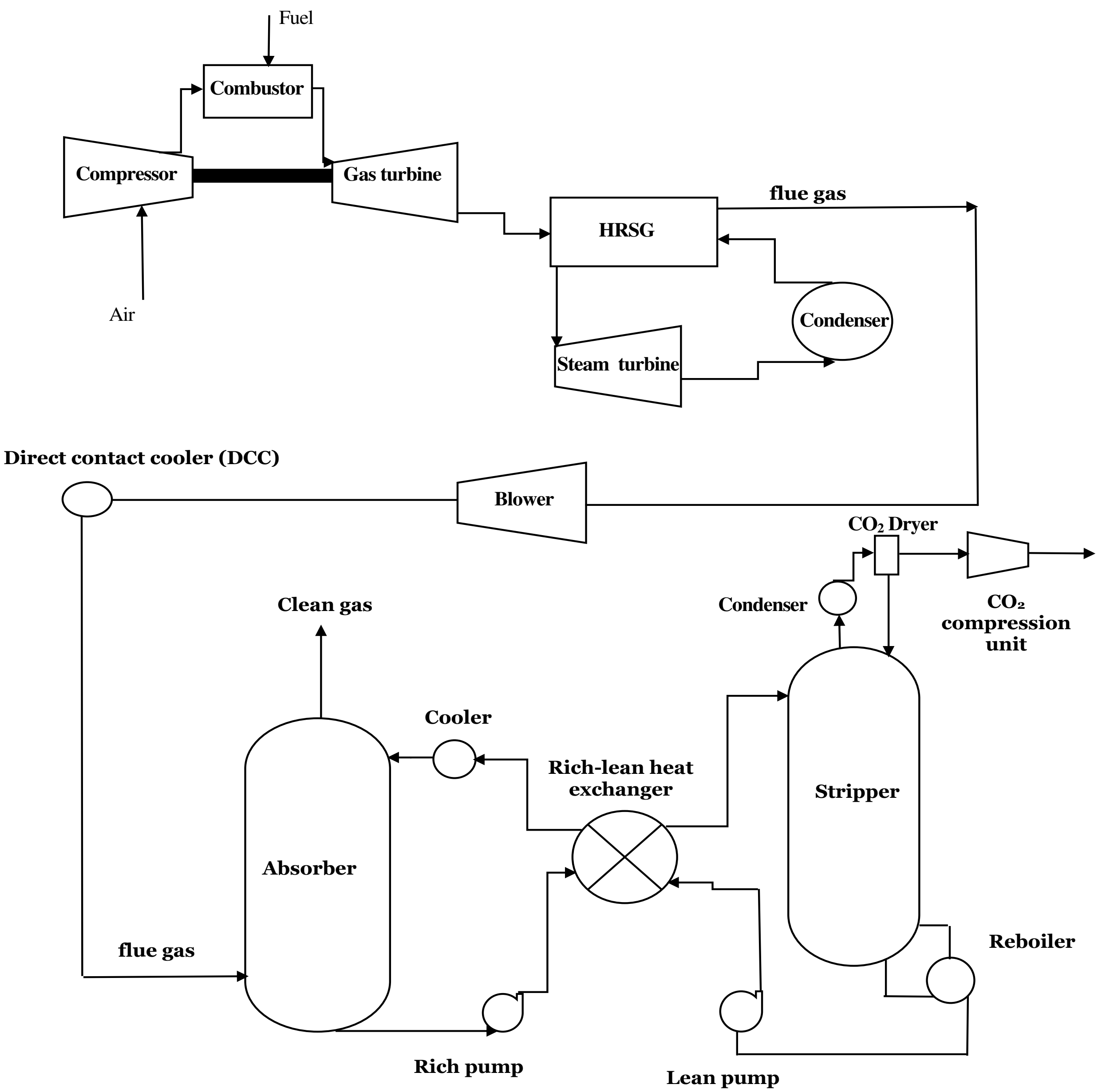

Fig 1. Flow diagram of the natural gas combined cycle plant integrated with a carbon capture plant

2.2 Assumptions for the NGCC plant and the PCC plant

Table 1 illustrates the assumptions for the NGCC plant, which was designed according to the DOE/NETL May 2023 report case CC2A-F [24]. Table 1 describes the details of the assumptions for the carbon capture plant, based on recommendations from the literature [6, 9, 25]. Table 2 shows the equilibrium and kinetic reactions employed in the absorber and stripper.

Table 1. Input assumptions for natural gas combined cycle and carbon capture plant

NGCC plant parameters	Value	Carbon capture plant parameters	Value
Pressure ratio	17	Thermodynamic model	ENRTL-RK
Compressor isentropic efficiency (%)	85	Column packing type	MellaPack 250Y
Inlet air flow rate (kg/h)	3805186	Capture efficiency (%)	90
Air inlet temperature (°C)	15	MEA concentration (%wt)	30
Inlet fuel flow rate (kg/h)	96104	Flue gas pressure at absorber inlet (bar)	1.11
Fuel inlet temperature (°C)	27	Flue gas temperature at absorber inlet (°C)	40
Fuel composition (vol%)		Lean solvent temperature at absorber inlet (°C)	40
CH_4	93.1	Stripper pressure (bar)	1.62
C_2H_6	3.2	Log mean temperature difference (LTMD) for rich-lean heat exchanger (°C)	8
C_3H_8	0.7	Stripper condenser temperature (°C)	40
C_4H_{10}	0.4	Rich/lean pump outlet pressure (bar)	3
CO_2	1.0	CO_2 compression pressure (bar)	152.7
N_2	1.6	Flooding capacity (%)	80
Fuel lower heating value (LHV) (MJ/kg)	47.2	Blower efficiency (%)	90
Turbine isentropic efficiency (%)	88		
Steam turbine isentropic efficiency ($HP^a/IP^b/LP^c$) (%)	92/ 92 /86		
$HP^a/IP^b/LP^c$ steam temperature (°C)	585/584/305		

HP^a - high pressure.
IP^b - intermediate pressure.
LP^c - low pressure.

Table 2. The equilibrium and kinetic reactions (from the Aspen Plus library)

Reaction Number	Reaction Type	Reactions
1	Equilibrium	$MEAH^+ + H_2O \leftrightarrow MEA + H_3O^+$
2	Equilibrium	$2H_2O \leftrightarrow H_3O^+ + OH^-$
3	Equilibrium	$HCO_3^- + H_2O \leftrightarrow CO_3^{-2} + H_3O^+$
4	Kinetic	$CO_2 + OH^- \rightarrow HCO_3^-$
5	Kinetic	$HCO_3^- \rightarrow OH^- + CO_2$
6	Kinetic	$MEA + CO_2 + H_2O \rightarrow MEACOO^- + H_3O^+$
7	Kinetic	$MEACOO^- + H_3O^+ \rightarrow MEA + CO_2 + H_2O$

2.3 EGR and SEGR assumptions

Fig 2 shows the schematic diagram for exhaust gas recirculation (EGR) and selective exhaust gas recirculation (SEGR). For EGR, 33% of the exhaust gas passes through the blower and is then cooled to 32°C [4]. After that, the cooled recycled exhaust gas is mixed with the air in the gas turbine compressor. The 33% EGR is the maximum value that can keep the oxygen level at the inlet combustor at 16% mol [5]. The mass flow rate of air for EGR was adjusted to keep the inlet gas turbine temperature at the design value.

For SEGR, this study adopts the design proposed by Merkel et al. [26] . In this configuration [6, 9, 26-27], the ambient air passes through the membrane system to separate CO_2 from other flue gases, including N_2 and O_2. In this design, two air blowers are required to overcome the pressure drop (10 kPa) on both the flue gas and air sides of the membrane system [26, 27]. The CO_2 permeance was 2200 GPU [6, 26, 27], while the cost of the membrane-installed skid was $50 per square meter [6, 26, 27].

According to Diego et al. [27], the optimal value to achieve a 90% capture rate is a 53% SEGR ratio with a CO_2 membrane separation efficiency of 95%. The SEGR cooled to 20 °C [27] before being mixed with air in the gas turbine compressor. The mass flow rate of air for SEGR was adjusted to maintain the inlet gas turbine temperature at the design value [27].

It should be noted that EGR and SEGR used a different cooler than DCC. This is because the DCC cools the flue gas to 40 °C [6, 9, 25], which is the standard value of flue gas entering the absorber. However, the EGR and SEGR require additional cooling below 40 °C to prevent a significant decline in the gas turbine power output [4].

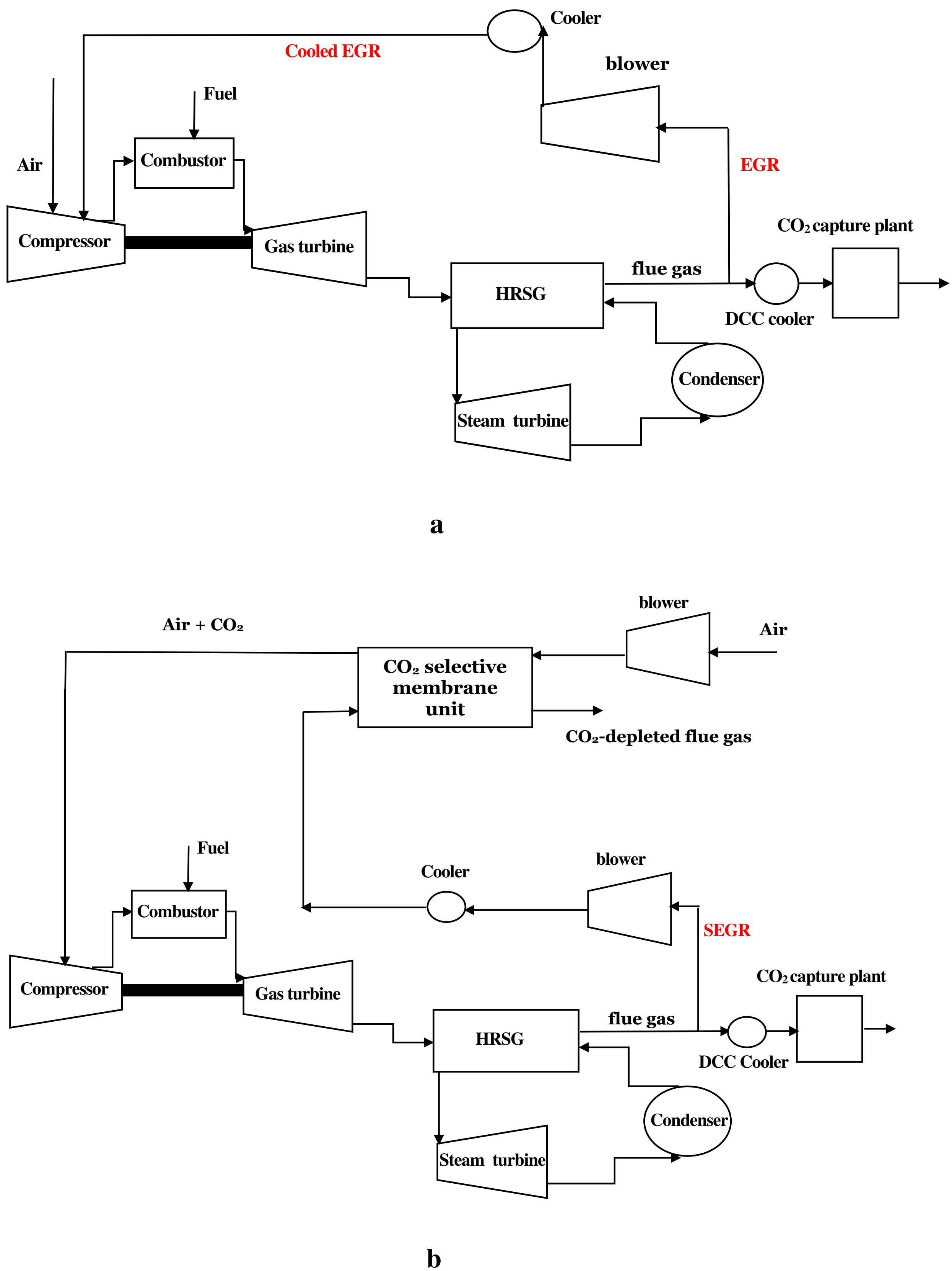

Fig 2. Process flow diagrams of (a) exhaust gas recirculation and (b) selective exhaust gas recirculation

2.4 Lean vapor compression (LVC) modification

As demonstrated in Fig. 3, this modification operates by flashing the hot lean solvent at atmospheric pressure to produce vapor and liquid phases. The vapor phase is compressed and fed to the bottom of a stripper, whereas the liquid phase is pumped to the rich-lean heat exchanger. The compressor isentropic efficiency is set at 90%. The compressor outlet pressure for LVC is fixed at 162 kPa, matching the stripper pressure to prevent solvent degradation. After applying LVC, the rich-lean heat exchanger's duty drops significantly because it only handles the liquid phase of the hot lean solvent (blue line). While the compressor handles the vapor phase. Furthermore, the heat duty of the rich-lean heat exchanger decreased because the reboiler duty decreased with LVC. Accordingly, this results in less heat needing to be transferred through the rich-lean heat exchanger.

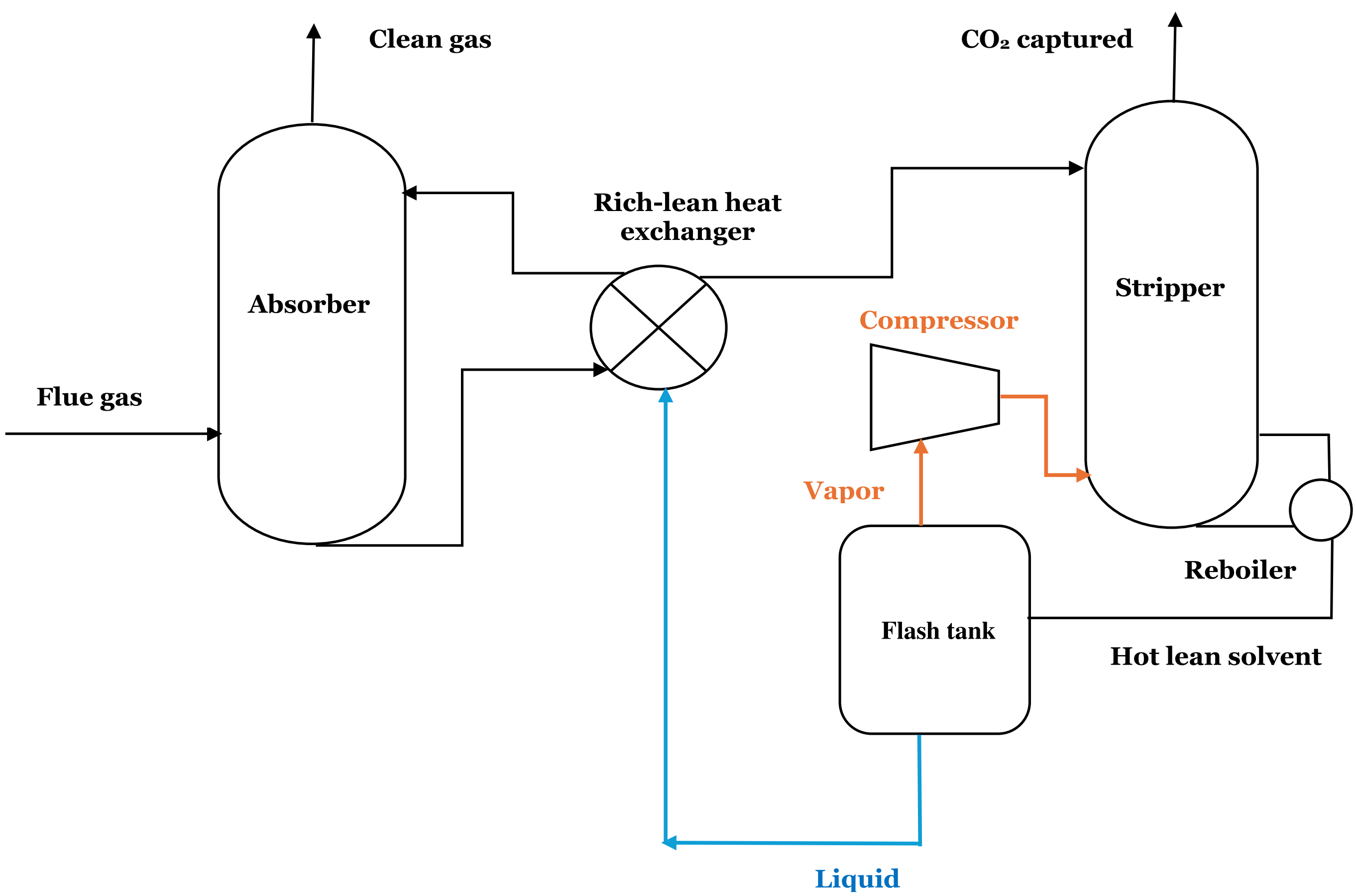

Fig. 3. Process flowsheet of the LVC modification

2.5 Economic model

The economic approach employed in this study to calculate the power plant cost is based on the guidelines published by DOE/NETL [28], which uses capital cost scaling. The capital cost scaling equation is expressed as follows:

$$SC = RC \times \left(\frac{SP}{RP}\right)^{Exp} \tag{1}$$

Where *SC* is the scaled cost, *RC* is the reference cost, *SP* is the scaling parameter, *RP* is the reference parameter, and *Exp* is the exponent.

Exponents are utilized to evaluate the cost of equipment and processes taken from the DOE/NETL report [28], as shown in the Table 3 . It is worth noting that the cost method recommended by the DOE/NETL for the carbon capture plant has certain limitations. This methodology relies solely on the mass flow rate of flue gas entering the absorber and the CO_2 product flow rate (see Table 3). This means that, with and without LVC, the mass flow rate of flue gas feed to the absorber and the CO_2 product flow rate remain unchanged, which makes the methodology suggested by DOE unsuitable for the PCC plant operating with process modifications. Therefore, for the PCC plant, the cost correlation methodology is used, as illustrated in Table 4. This methodology is widely used in some studies [6,29] to calculate the cost of the PCC plant. To account for inflation, the capital cost of the equipment for the carbon capture plant was adjusted to US$2024 using the Chemical Engineering Plant Cost Index (CEPCI). The cost update was performed using the following equation:

$$C_{2024} = C_{ref} \times \frac{CEPCI_{2024}}{CEPCI_{ref}} \tag{2}$$

Where C_{2024} is the updated cost in 2024, C_{ref} is the reference cost from the original year, and $CEPCI_{2024}$ and $CEPCI_{ref}$ are the cost factors for 2024 and the reference year, respectively.

Table 3. Scaling parameters suggested by DOE/NETL [28]

Process area	Parameter description	Scaling exponent
Feed water system	HP feed water flow rate	0.72
EGR system	EGR flow rate	0.70
HRSG system	HRSG duty	0.70
Steam turbine system	Steam turbine power	0.80
Cansolv CO_2 Removal System	CO_2 Product Flow Rate, / Gas Flow to CO_2 Absorber	0.60 [A]
CO_2 Compression & Drying	CO_2 Product Flow Rate	0.77

[A] To scale the Cansolv CO_2 Removal System, 40% of the cost is scaled using the parameter Gas Flow to CO_2 Absorber; the remaining 60% is scaled using the parameter CO_2 Product Flow Rate.

Table 4. Cost correlations for carbon capture plant equipment

Equipment type	Costing parameter	Equation
Absorber and stripper (USD 2001) [30]	Diameter, D (m), Height H (m)	$H \times (10^{0.563\,(\log D)^2 + 1.056\,(\log D) + 3.8057})$
Column packing (USD 2009) [31]	Packed volume, V (m^3)	$1700 \times V$
Heat exchanger (USD 2000) [32]	Area, A (m^2)	$94093 + 1127 \times A^{0.98}$
Pump (USD 2001) [30]	Flow rate, F (m^3/h)	$10^{0.2468\,(\log F)^2 - 0.5966\,\log F + 3.9213}$
Compressor (USD 2002) [33]	Power, P (KWe)	$10^{0.9518\,\log P + 2.9184}$

3. Model validation

3.1 Power plant model validation

The model was validated with the DOE/NETL May 2023 report, case CC2A-F [24], which provides details about partial load conditions for the NGCC plant. The mass flow rate of fuel for partial loads was directly taken from the NETL report. The mass flow rate of air was adjusted to match the gas turbine power output reported by the NETL report. Similarly, the steam flow rate was adjusted to match the steam turbine's power output, as outlined in the NETL report. As shown in Table 5, the model demonstrates excellent agreement with the NETL report across all gas turbine loads.

Table 5 . Model validation results for the NGGC plant with the NETL report

Gas turbine loads (%)	100		80		60	
	NETL	Model	NETL	Model	NETL	Model
Mass flow rate of fuel (kg/h)	96104		77629		63366	
Gas turbine power output (MWe)	486	485.4	389	391.1	292	293.7
Steam turbine power output (MWe)	265	267.6	210	210.5	185	184.6
Total Auxiliaries (MWe)	12	12.8	9	10.1	8	8.6
Total net power (MWe)	738	740.2	589	591.5	469	469.7
Thermal efficiency (%)	58.6	58.7	57.9	58.1	56.4	56.5
Flue gas flow rate (kg/s)	1083.7	1083.7	-	901.5	-	747.6
CO_2 in the exhaust gas (mol%)	4.23	4.23	-	4.10	-	4.04

3.2 Carbon capture model validation

The carbon capture plant model is validated using the experimental data obtained from Notz et al. [34], who performed 47 experiments. Seven cases from 47 experiments were selected for model validation. These selected cases were chosen because they exhibit both high and low CO_2 levels in flue gas, which simulates the behavior of baseline, EGR, and SEGR. Fig. 4 present the validation results, which show an average error of less than 5%, indicating a high level of confidence in the model's accuracy.

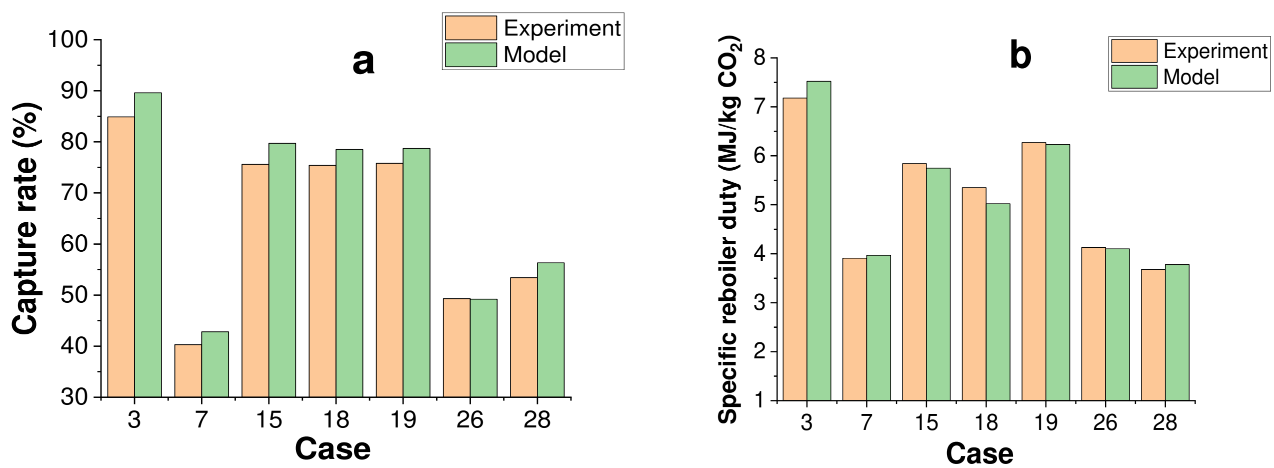

Fig. 4. Model validation for (a) capture rate, and (b) reboiler duty

4. Results and discussion

4.1 Effect of lean loading on specific reboiler duty

Determining the optimal lean loading is essential to minimizing solvent regeneration duty for the PCC plant. Fig. 5 presents the impact of lean loading from 0.14 to 0.28 mol CO_2/mol MEA on the solvent regeneration heat duty for different configurations.

Examining Fig. 5 in more detail, it is evident that all proposed configurations achieved their optimal lean loading approximately at 0.21, where the reboiler duty reaches its minimum value. Although the EGR and SEGR configurations increase CO_2 content in the flue gas, the fundamentals of thermodynamics governing MEA absorption and desorption remain unchanged.

This explains why all configurations exhibited the same lean loading curve. LVC can perform optimally under all lean loading and configurations tested, offering significant energy savings. SEGR with LVC achieves minimum solvent regeneration duty at lean loadings of 0.14 to 0.24. However, at a lean loading of 0.28, SEGR has the highest reboiler duty due to the high CO_2 content in the flue gas and high lean loading.

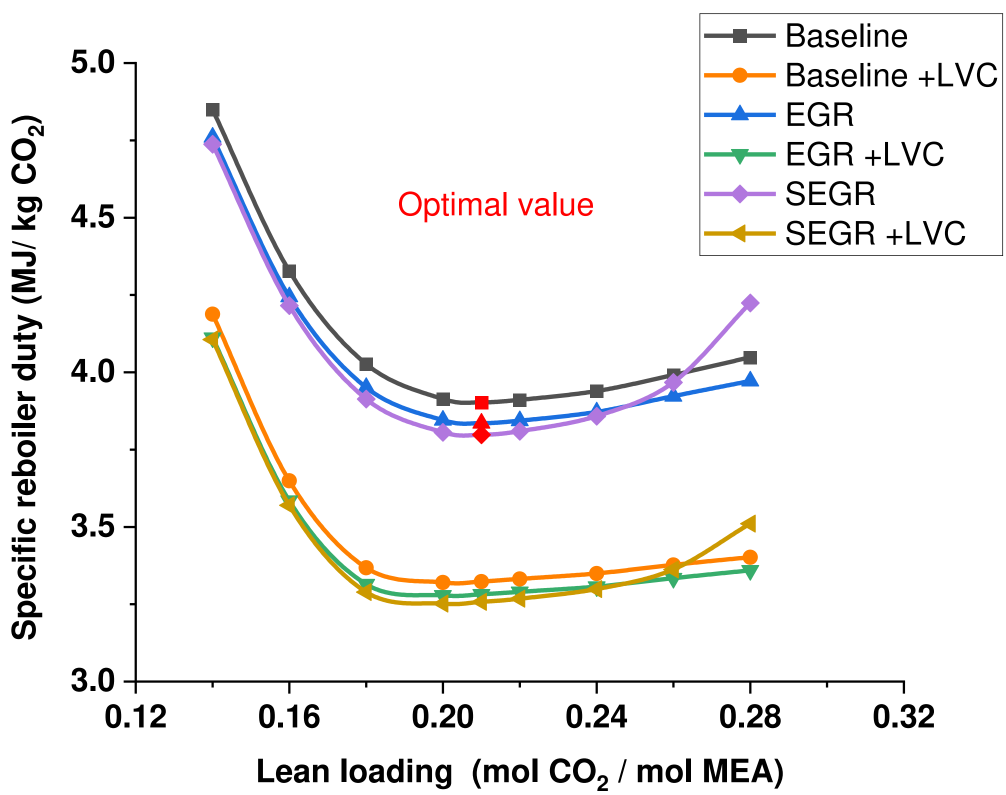

Fig. 5. Impact of lean loading on the specific reboiler duty

4.2 Absorber optimization

Optimizing the absorber height plays a key role in minimizing the capital cost for the PCC plant. Fig 6 illustrates how a change in absorber height, ranging from 8 m to 30 m, influences reboiler duty across different scenarios. The capture rate was fixed at 90% for all heights assessed, while the solvent flow rate was adjusted for each height to achieve the target capture rate. The optimal height was determined when the increase in absorber height resulted in a negligible reduction in reboiler duty and L/G ratio.

Fig 6 demonstrated that increasing the absorber height significantly reduces the reboiler duty for all cases, particularly from 8 m to 13 m. However, beyond 13 m, a slight reduction in SRD can be observed for all scenarios. Accordingly, the optimal height for the baseline is 17 m, while the EGR and SEGR are 15 m and 16 m, respectively. The main reason for the minor difference in optimal absorber height between baseline, EGR, and SEGR is that the amount of CO_2 captured is the same in all configurations. The SEGR has a slightly higher optimal height than the EGR due to the high CO_2 concentration in the flue gas, which requires a larger height. With LVC, for every configuration, 1 m of saving in absorber height can be gained.

It should be noted that the considerable difference between these configurations is in the absorber diameter. For example, the diameters needed to meet the flooding capacity of 80% for the baseline, EGR, and SEGR are 21.3 m, 18.6 m, and 16.4 m, respectively.

Fig 7 shows the influence of absorber height on the L/G ratio. After a certain height, the increase in absorber height results in a negligible reduction in the L/G ratio. As a result, the optimal L/G ratios for baseline, EGR, and SEGR are 1.02, 1.5, and 2.13, respectively. These values correspond to the optimal absorber height mentioned above. It is essential to note that the EGR and SEGR have a higher L/G ratio than the baseline due to the higher CO_2 concentration in the flue gas.

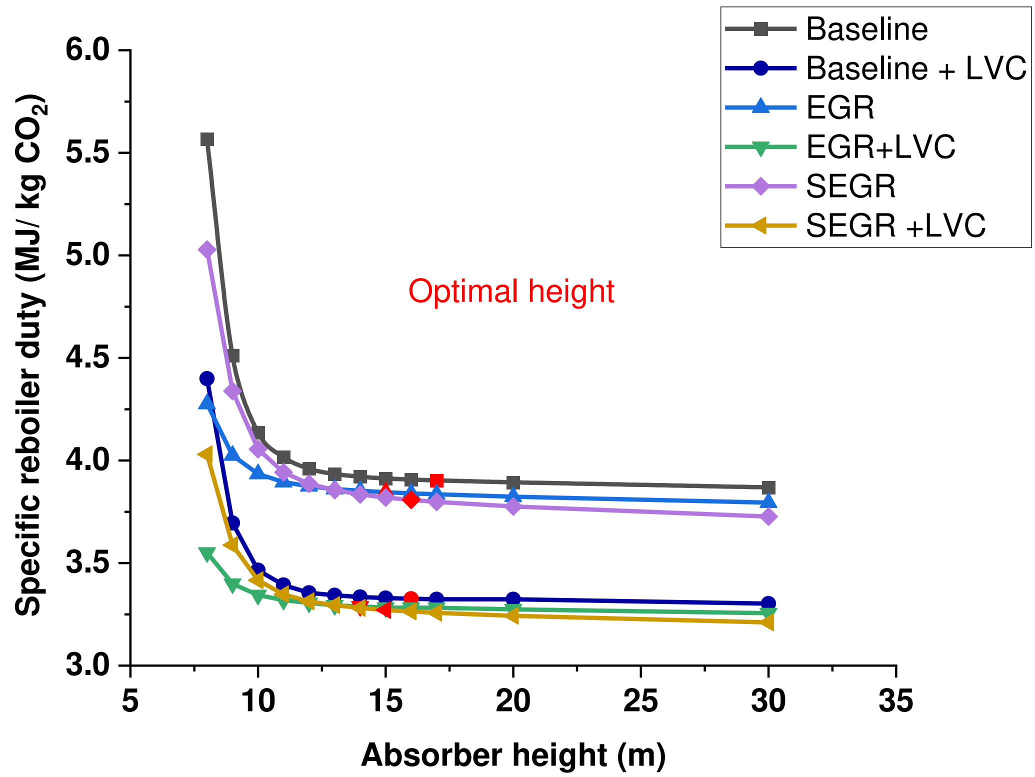

Fig 6. Impact of absorber height on the specific reboiler duty

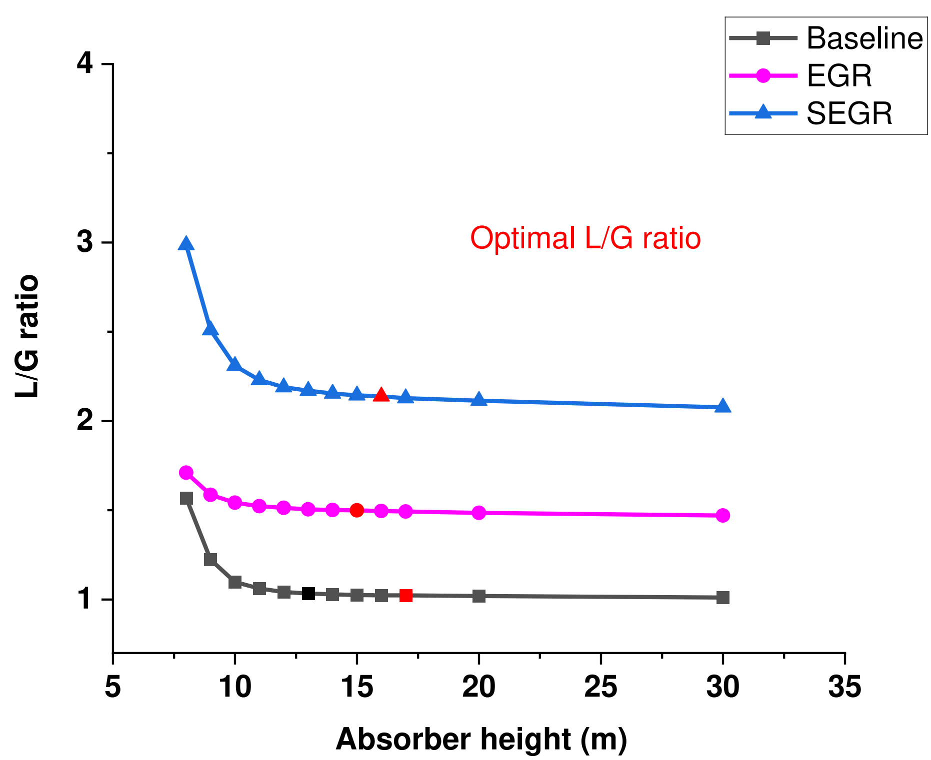

Fig 7. Impact of absorber height on the L/G ratio

4.3 Stripper optimization

A sensitivity analysis was conducted to examine the effect of varying the stripper height, ranging from 5 m to 30 m, on the reboiler duty. In this analysis, only the stripper height was adjusted, while other parameters were constant.

Fig. 8 reveals a similar pattern to the absorber height, where marginal energy savings are gained after a certain stripper height. For instance, minor energy savings are achieved after 17 m. Accordingly, the optimal stripper height for the baseline is 16 m, whereas the EGR and SEGR are 15 m. It was noted that integrating LVC to the PCC plant results in 4 m of savings in the stripper height for every configuration.

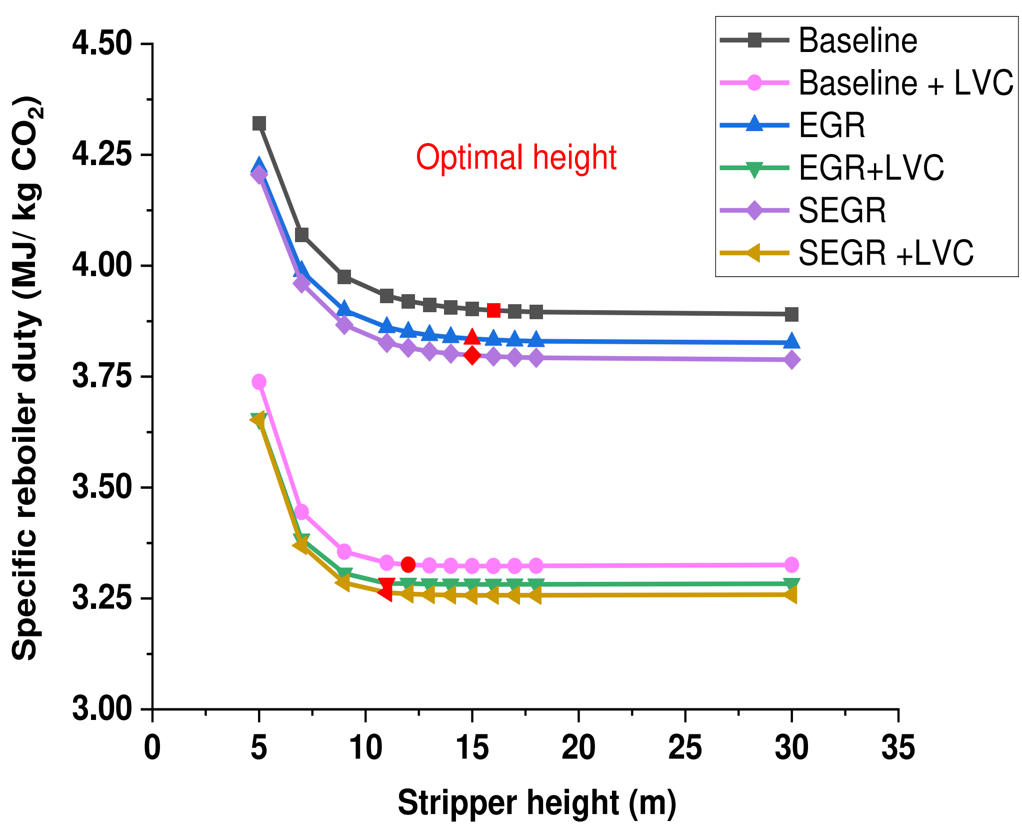

Fig. 8. Impact of stripper height on the specific reboiler duty

4.4 Effect of the LVC on the baseline performance

Table 6 illustrates the influence of the gas turbine (GT) loading, ranging from 100% to 60%, on the baseline configuration with and without LVC. The results are compared to the NETL report case B31B.90 [4]. Slight differences are observed in parameters such as gas turbine power output, flue gas properties, and total auxiliary power consumption compared to the NTEL report. This is because, as mentioned before, this paper used a 751 MW plant, whereas B31B.90 used a 740 MW plant (see methodology section).

The steam turbine power output after the PCC plant integration is slightly lower in this study (202.8 MW) compared to the NETL report (215 MW). This is because NETL employed an advanced solvent named Shell's amine solvent-based CANSOLV system for the carbon capture plant. This results in a lower steam flow rate extracted from the steam turbine for the reboiler. In contrast, this investigation employed MEA solvent.

It is clear from the table below that the specific reboiler duty experiences a slight decline under partial load conditions. This is because under partial loads, the flue gas mass flow rate decreases, which can contribute to minimizing the SRD. However, when the GT load decreases, the mole fraction of CO_2 in the flue gas decreases, which requires higher solvent regeneration duty to achieve a 90% removal rate. These opposing conditions in the exhaust gas result in a relatively slight change in SRD.

The mass flow rate of steam for the reboiler is extracted from the intermediate pressure to low-pressure (IP/LP) crossover [6, 35] at a pressure of 510 kPa. This pressure exceeds the reboiler need, so it must be reduced to 250 kPa before entering the reboiler. At 250 kPa, the saturated steam temperature is 127.4 °C (according to the steam table). The reboiler in this work operates at 117.5 °C for all configurations. Consequently, this ensures a 10 °C temperature difference (127.4 - 117.5) between the steam and reboiler to prevent solvent degradation [6, 35]. To reduce the quantity of mass flow rate extracted for the reboiler, condensed steam in the reboiler is returned to the HRSG [6, 35].

According to Table 6, introducing LVC reduced the steam mass flow rate extracted for the reboiler for all GT loads. It is worth noting that the LVC compressor consumes a small amount of power (2.4 MW) because it slightly raises the pressure of the hot lean vapor from 101 to 162 kPa. This finding aligns with a study by Asadi et al. [18], which found that LVC can reduce the reboiler duty by 13%, with compressor power consumption of 2.4 MW. Consequently, LVC boosts the steam turbine power output from 8 MW to 5 MW as the GT load varied from 100% to 60%. This improvement contributes to a 0.4 percentage point increase in thermal efficiency.

Table 6. Performance of the baseline with and without LVC under partial loads

Gas turbine loading (%)	100 B31B.90 [4]	100	80	60
Mass flow rate of flue gas (kg/s)	1090.9	1083.7	901.5	747.6
CO_2 concentration of flue gas (mol%)	4.08	4.23	4.10	4.04
CO_2 captured (kg/s)	62.2	65.0	52.5	42.9
Rich loading	-	0.481	0.481	0.481
L/G ratio (kg/kg)	-	1.02	0.99	0.97
Specific reboiler duty (MJ/kg CO_2)	-	3.90	3.90	3.89
Steam flow rate extracted for CO_2 capture (kg/s)	-	101.1	81.6	66.5
Gas turbine power output (MWe)	477	485.4	391.1	293.7
Steam turbine power output without mass flow rate extracted (MWe)	263	267.6	210.5	184.6
Steam turbine power output with mass flow rate extracted (MWe)	215	202.8	158.2	141.9
Auxiliary power consumption (MWe) (power plant + capture plant + CO_2 compression unit)	47	57.4	46.9	40.7
Net power plant power output (MWe)	645	630.8	502.4	394.9
Thermal efficiency (%)	52.7	50.06	49.36	47.53
Lean vapor compression				
Specific reboiler duty (MJ/kg CO_2)	-	3.40	3.40	3.40
Lean vapor compressor (MW)	-	2.4	1.9	1.6
Auxiliary power consumption (MWe)	-	59.8	48.8	42.3
Steam flow rate extracted for CO_2 capture (kg/s)	-	88.2	71.3	58.2
Steam turbine power output (MWe)	-	211.1	164.8	147.3
Net power plant power output (MWe)	-	636.7	507.1	398.7
Thermal efficiency (%)	-	50.53	49.82	47.99

4.5 Effect of the LVC on the EGR performance

Implementing EGR reduces the flue gas feed to the carbon capture unit by up to 33% and increases the CO_2 concentration in flue gas to approximately 1.5 times the baseline concentration. As demonstrated in Table 7, employing the EGR results in higher rich loading and L/G than the baseline due to a higher CO_2 level in flue gas.

Implementing EGR improves the NGCC plant efficiency by reducing steam extraction for the reboiler and lowering the PCC plant's auxiliary power. This reduction occurs because EGR leads to a decrease in the mass flow rate of flue gas feed to the carbon capture plant.

Additionally, after applying EGR, the exhaust gas temperature of the gas turbine [6, 27] increased compared to the baseline due to changes in the air inlet properties, including molecular weight, density, and specific heat. Accordingly, higher exhaust gas turbine temperatures mean more heat is available for the steam cycle, increasing the power output of the steam cycle. However, the gas turbine power output slightly decreases because the recycled exhaust enters at 32 °C [4], which is above the ambient temperature at which the air enters the gas turbine compressor.

According to Table 7, EGR could increase the thermal efficiency by up to 0.3% points compared to the baseline under full and partial loads. It is worth noting that the EGR+LCV has the highest thermal efficiency, increasing it by 0.7 % points compared to the baseline.

Table 7. Performance of the EGR with and without LVC under partial loads

Gas turbine loading (%)	100	80	60
Mass flow rate of flue gas (kg/s)	724.9	603.1	500.3
CO_2 concentration of flue gas (mol%)	6.32	6.14	6.04
CO_2 captured (kg/s)	64.9	52.4	42.7
Rich loading	0.487	0.487	0.487
L/G ratio (kg/kg)	1.49	1.44	1.42
Specific reboiler duty (MJ/kg CO_2)	3.83	3.82	3.82
Steam flow rate extracted for CO_2 capture (kg/s)	99.3	80.0	65.1
Gas turbine power output (MWe)	480.4	387	290.4
Steam turbine power output without mass flow rate extracted (MWe)	272.5	214.4	187.7
Steam turbine power output with mass flow rate extracted (MWe)	208.8	163	145.9
Auxiliary power consumption (MWe) (power plant + capture plant + CO_2 compression unit+ EGR)	54.5	44.4	38.8
Net power plant power output (MWe)	634.7	505.6	397.5
Thermal efficiency (%)	50.37	49.67	47.84
Lean vapor compression			
Specific reboiler duty (MJ/kg CO_2)	3.36	3.35	3.35
Lean vapor compressor (MW)	2.3	1.8	1.5
Auxiliary power consumption (MWe)	56.8	46.2	40.3
Steam flow rate extracted for CO_2 capture (kg/s)	87.0	70.1	57.2
Steam turbine power output (MWe)	216.7	169.3	150.9
Net power plant power output (MWe)	640.3	510.1	401
Thermal efficiency (%)	50.81	50.11	48.26

4.6 Effect of the LVC on the SEGR performance

The SEGR configuration minimizes the exhaust gas volume fed to the MEA-based CO_2 capture plant by up to 53 % and increases the CO_2 content in exhaust gas by up to 2.1 times the baseline level, as illustrated in Table 8. The SEGR has a higher rich loading and L/G ratio than the EGR and baseline because it has a higher CO_2 level in the flue gas. The lowest solvent regeneration duty was reported with the SEGR and LVC.

As with EGR after applying SEGR, the mass flow rate of steam extracted for the reboiler decreased, and the exhaust gas turbine temperature increased. This results in more heat available for the steam turbine cycle.

The SEGR scenario has higher total power consumption than the EGR and baseline. This is mainly due to the SEGR requiring significant air blower power to overcome the membrane pressure drop (see Fig 2). As a result, SEGR's thermal efficiency is lower than that of the baseline and EGR. These results are consistent with the literature [27]. The LVC boosts the thermal efficiency of SEGR by 0.4 % points.

Table 8. Performance of the SEGR with and without LVC under partial loads

GT load (%)	100	80	60
Mass flow rate of flue gas (kg/s)	502.3	417.7	346
CO_2 concentration of flue gas (mol%)	8.83	8.57	8.45
CO_2 captured (kg/s)	64.9	52.4	42.8
Rich loading	0.490	0.491	0.492
L/G ratio (kg/kg)	2.13	2.05	2.02
Specific reboiler duty (MJ/kg CO_2)	3.8	3.78	3.77
Steam flow rate extracted for CO_2 capture reboiler (kg/s)	98.3	79.0	64.3
Gas turbine power output (MWe)	479.1	386.2	290.6
Steam turbine power output without mass flow rate extracted (MWe)	271.4	213.5	186.7
Steam turbine power output with mass flow rate extracted (MWe)	208.4	162.9	145.4
Auxiliary power consumption (MWe) (power plant + capture plant + CO_2 compression unit + SEGR)	66.8	54.7	48
Net power plant power output (MWe)	620.7	494.4	388.0
Thermal efficiency (%)	49.26	48.57	46.70
Lean vapor compression			
Specific reboiler duty (MJ/kg CO_2)	3.33	3.32	3.31
Lean vapor compressor (MW)	2.3	1.8	1.5
Auxiliary power consumption (MWe)	69.1	56.5	49.5
Steam flow rate extracted for CO_2 capture (kg/s)	86.3	69.5	56.6
Steam turbine power output (MWe)	216	169.1	151
Net power plant power output (MWe)	626.0	498.8	392.1
Thermal efficiency (%)	49.68	49.00	47.19

5. Economic analysis and results

In this section, all proposed configurations are evaluated economically, as shown in Table 9 . The cost was compared with the DOE/NETL case B31B.90 [4]. It is worth mentioning that the scaling parameters methodology suggested by the NETL report is not appropriate for the carbon capture plant with process modifications [29]. This is because the NETL recommended calculating the cost of a carbon capture plant based on the mass flow rate of exhaust gas feed to the absorber and the CO_2 product flow rate (see Table 3). Consequently, this means that with and without LVC, the mass flow rate of flue gas feed to the absorber and the CO_2 product flow rate are the same. Therefore, the cost correlations methodology (see Table 4) was employed, which is commonly used in many studies in the literature [6,29] to calculate the cost of the PCC plant.

According to Table 9, the use of the scaling parameters methodology yields cost results that closely align with those reported by the NETL report. For example, according to the NETL report, the cost of a carbon capture plant was \$477.7 M, whereas our model estimated it to be \$ 485.7 M.

In contrast, when the cost correlations method is used, a significant difference is observed between our results and those of the NETL report. For example, the PCC plant cost according to the NETL was \$477.7 M, whereas our model in the baseline was \$286.2 M. This is because the cost correlations method depends on the PCC plant equipment specifications, such as the size of the absorber and the area of the heat exchanger (see Table 4). The NETL report does not provide details of the height, diameter, and packing materials of the absorber and stripper. Also, the NETL report does not provide information about the assumptions made about the heat exchangers for the PCC plant. Additionally, the NETL employed an advanced solvent called Shell's amine solvent-based CANSOLV system, whereas this work utilized MEA. Therefore, it's challenging to replicate the cost results reported by the NETL report using the cost correlations method.

According to Table 9, implementing EGR and SEGR reduces the cost of the PCC plant by 10% and 16 % respectively, compared to the baseline. The reason for this reduction is that the EGR and SEGR reduce the mass flow rate of flue gas feed to the carbon capture plant, thereby minimizing the size of the PCC plant equipment.

The cost of CO_2 compression and drying is the same for all cases because the amount of CO_2 captured is fixed. The cost of the HRSG and steam turbine for EGR and SEGR is higher than the baseline due to the lower mass flow extracted for the reboiler, resulting in a higher steam power output. The gas turbine cost was the same for all cases because the same gas turbine was used for all cases.

According to Shirdel et al. [36], the rich lean heat exchanger accounts for 29 % of the total cost of the carbon capture plant. As shown in the Table 9, the rich-lean heat exchanger represents between 20% and 30% of the total cost of the PCC plant in all cases. Applying LVC can remarkably reduce the size and cost of the rich-lean heat exchanger because it handles only the liquid phase of the hot lean solvent (see Fig. 3). In addition, with LVC, the reboiler duty decreases, resulting in less heat being transferred through the rich-lean heat exchanger. Consequently, with LVC, the cost of the lean-rich heat exchanger decreased by approximately 30% for all configurations. This represents the major contributor to the cost reduction of the PCC plant with LVC.

With LVC, the cost of the stripper condenser is also reduced because less vapor is treated at the top of the stripper. Furthermore, with LVC, the height of the stripper and absorber decreased as discussed in the optimization section for the absorber and stripper. It is worth noting that the diameter of the stripper with LVC increases slightly because the height of the stripper decreases significantly. The lean vapor compressor incurs a minor cost because it increases the pressure of the hot lean solvent vapor slightly from 101 kPa to 162 kPa. This increase in pressure is not substantial. Similarly, the flash tank of LVC also has a minor cost.

The total plant cost (TPC) is calculated by summing the costs of the PCC and the NGCC plant. The total overnight cost (TOC) is calculated by summing the TPC with additional expenses, including pre-production, inventory capital, and other costs. The SEGR has the highest TPC and TOC since it involves a membrane system. This system includes the membrane device, two air blowers, and a cooler (see Fig 2). These findings align with the literature [6, 27], which has shown that SEGR has higher TPC and TOC than EGR and the baseline. The EGR with LVC has the lowest TPC and TOC.

Our results align with those reported by Arshad et al. [15], which indicate a 5.81% reduction in cost for the NGCC with PCC plant using LVC. Similarly, a study by Yagihara et al. [37] observed a 10% decrease in PCC plant equipment cost, from \$14.7M to \$13.3M, for an NGCC integrated with an MEA-based PCC plant using LVC. Both studies [15,37] attribute the cost reduction primarily to significant cost savings in the rich-lean heat exchanger, which outweigh the additional costs of the lean vapor compressor and flash tank. Our cost results for baseline showed an 11% reduction in PCC plant equipment cost with LVC. The primary reason our cost is slightly higher than that in the literature [15,37] is that we optimized the stripper and absorber height with and without LVC, which neither of the two studies did. Additionally, our cost calculation methodology, power plant capacity, and assumptions for the PCC plant differ from those in the literature [15,37].

Table 9. Cost analysis for all configurations

Methodology	Scaling parameters (NETL)		Cost correlation					
	B31B.90 (NETL) [4]	Our model	Baseline	Baseline +LVC	EGR	EGR+LVC	SEGR	SEGR+LVC
Absorber height (m)	-	-	17	16	15	14	16	15
Absorber diameter (m)	-	-	21.3	21.3	18.6	18.6	16.4	16.4
Stripper height (m)	-	-	16	12	15	11	15	11
Stripper diameter (m)	-	-	9.3	9.6	9.3	9.6	9.2	9.5
Rich lean heat exchanger duty (MW)	-	-	259.8	200.6	254.0	195.6	243.8	186.5
Cost analysis results (M$ 2024)								
Absorber	-	-	62.8	59.1	40.3	37.6	32.2	30.2
Stripper	-	-	9.4	7.5	8.8	6.9	8.6	6.8
Reboiler	-	-	45.8	40.1	45.2	39.7	44.7	39.4
Rich lean heat exchanger	-	-	75.4	53.4	74.7	52.2	71.7	49.8
Stripper Condenser	-	-	5.1	3.3	5.0	3.2	5.0	3.2
Lean vapor compressor	-	-	-	2.7	-	2.7	-	2.7
Flash tank for LVC	-	-	-	1.04	-	1.04	-	1.04
Pumps	-	-	0.34	0.34	0.33	0.33	0.33	0.33
Lean cooler	-	-	4.20	4.20	4.20	4.20	4.20	4.20
DCC cooler	-	-	6.4	6.4	5.0	5.0	3.2	3.2
Exhaust gas air blower	-	-	12.3	12.3	8.40	8.40	5.8	5.8
CANSOLV Carbon Dioxide (CO_2) Removal System	415.2	421.3	221.9	190.4	191.9	161.3	175.7	146.7
CO_2 Compression & Drying	60.8	62.8	62.8	62.8	62.8	62.8	62.8	62.8
CO_2 Compressor Aftercooler	0.5	0.5	0.5	0.5	0.5	0.5	0.5	0.5
Gas Cleanup Foundations	1.1	1.1	1.1	1.1	1.1	1.1	1.1	1.1
Total carbon capture plant cost	477.7	485.7	286.2	254.9	256.4	225.8	240.2	211.1
Feed water system and natural gas pipeline	111.9	113.0	113.0	113.0	114.6	114.6	114.2	114.2
Cooling water system	50.6	50.6	50.6	50.6	51.1	51.1	51.7	51.7
EGR system	-	-	-	-	19.6	19.6	-	-
Membrane system	-	-	-	-	-	-	45.3	45.3
Gas turbine system	113.76	115.35	115.35	115.35	115.35	115.35	115.35	115.35
HRSG system	110.15	110.3	110.3	110.3	111.8	111.8	111.4	111.4
Steam turbine system	82.5	78.7	78.7	81.3	80.6	83.0	80.5	82.8
Instrumentation, control, buildings & structures, Accessory electric plant & improvements to site (13 % of TPC) *a	140.2	143.0	112.7	108.4	112	107	113.3	109.3
Total plant cost (TPC)	1087	1096	866	833	861	829	872	841
Pre-Production Costs (3.2 % of TPC) *b	34.26	35.07	27.7	26.6	27.5	26.5	27.9	26.9
Inventory Capital (0.6 % of TPC) *c	6.4	6.5	5.2	5	5.1	5	5.2	5
Other Costs (17.8 % of TPC) *d	193	195	154	148	153	147	155	149
Total overnight cost (TOC)	1321	1332	1053	1012	1047	1008	1060	1022

*a, *b, *c, and *d. These percentages are calculated by the authors based on the NETL report.

6. Conclusion

The present paper provides a techno-economic assessment of an innovative design that combines lean vapor compression (LVC) with the PCC plant and the NGCC plant, incorporating exhaust gas recirculation (EGR) and selective exhaust gas recirculation (SEGR). The main objective of this study was to evaluate the impact of LVC on the energy performance and economic viability of the baseline, EGR, and SEGR. The simulation findings showed that implementing 33% EGR increased the CO_2 concentration in the exhaust gas from a baseline of 4.2 mol% to 6.3 mol%, while 53% SEGR further elevated the CO_2 level to 8.8 mol%. Among all configurations assessed, SEGR+LVC achieved a 14 % reduction in solvent regeneration duty compared to the reference case, representing the highest reduction compared to other configurations. In contrast, the highest thermal efficiency was achieved with EGR + LVC, which ranged from 50.81% to 48.26% as the gas turbine (GT) loading varied from 100% to 60%. By comparison, the conventional NGCC + PCC plant had a thermal efficiency of 50.06% to 47.53 % over the same GT loading. The most significant capital cost reduction in the PCC plant equipment was achieved with SEGR+LVC, lowering the cost from the baseline value of $ 286 M to $211 M. However, the SEGR without LVC exhibited the highest total plant cost and total overnight cost of $872 M and $1,060 M, respectively, due to the additional costs for the CO_2 membrane system. The novel configurations proposed in this investigation contribute to the advancement of carbon capture research and offer valuable insights for industry decision-makers.

CRediT authorship contribution statement

Fayez Alruwaili: Writing – original draft, Validation, Methodology, Investigation, Formal analysis, Data curation, Conceptualization. **Kevin J. Hughes**: Writing – review & editing, Supervision, Methodology, Conceptualization. **Derek B. Ingham**: Writing – review & editing, Supervision. **Lin Ma**: Supervision, Conceptualization. **Mohamed Pourkashanian**: Supervision, Project administration, Conceptualization.

Declaration of competing interest

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

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