IECM Technical Documentation:

Amine-based Post-Combustion Process for Deep Carbon Capture

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Amine-based Post-Combustion Process for Deep Carbon Capture

Prepared

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Table of Contents

List of Figures

List of Tables

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1. Introduction

Most current economic and social activities heavily rely on energy supplies. Globally, primary energy sources for electricity generation are fossil fuels, while renewable energy sources are environmentally friendly. Renewable energy sectors are not entirely developed and need more improvements. In contrast, coal used for power generation is relatively low-cost, reliable, technically mature, and abundantly available in various regions of the world.

Regardless of the convenience of coal, it leads significantly to environmental degradation. The combustion of coal emits significant amounts of greenhouse gases, especially CO_2, into the environment. Climate change, primarily driven by CO_2 emissions, has led to a range of negative impacts, including global temperature rise, health issues, increased frequency and intensity of extreme weather events, and disruptions to global food security. To cope with this issue, a variety of low-carbon energy and environmental control technologies are required in order to meet the U.S. climate target for a carbon-free electric power sector by 2035 and a net-zero emissions economy by 2050 [1]. These technologies include renewables, battery energy storage, carbon capture and storage (CCS), bioenergy with CCS (BECCS), direct air capture and storage (DACS), and small modular nuclear reactors. Fossil fuels, which today supply around 60% of U.S. power and nearly half of the predicted generation needs in 2050 under existing policies [1], are regarded as critical to ensuring dependability, affordability, and operational flexibility in a system with significant variable-input renewable sources [2]. Therefore, to reach the emission reduction targets, the deployment of CCS will be required to decarbonize fossil fuel-based power plants, industrial processes of steel, cement, chemicals, and hydrogen production from fossil fuel resources.

Carbon capture can be adopted to the industrial and power sectors through the technologies of post-, pre-, or oxy-combustion CO_2 capture. Among these techniques, post-combustion CO_2 capture can be retrofitted easily to existing power plants and be integrated into newly built power plants. Post-combustion carbon capture is a mature and commercial technology [3]. It is in the category of chemical solvent-based absorption, membranes, cryogenic separation, and adsorption [4]. Among them, chemical solvent-based absorption is a feasible way to abate CO_2 emissions [5]. Chemical solvents are predominantly amines, including monoethanolamine (MEA), 2-amino-2-methyl-1-propanol (AMP), N-methyldiethanolamine (MDEA), diethanolamine (DEA), and their mixtures. Among these amines, MEA has the highest absorption power, with the ranking order of MEA > AMP > DEA ≫ MDEA [6].

Until date, CCS designs have been based on a 90% CO_2 capture rate, regarded as the most cost-effective option [7]. Following a surge of activity two decades ago, interest and investment in CCS declined due to projected climate change policy drivers producing markets for CCS. The recent adoption of climate targets that require net-zero emissions has reignited interest in CCS as a vital technology for the industrial and electric power sectors. Limited recent research on coal- and gas-combustion power plants suggests that increasing CO_2 removal efficiency from 90% to 99% or more can achieve net-zero emissions with rising CO_2 avoidance costs only by 3–13% or less [7, 8]. Carbon capture rates ranging from 95–99%, referred to as 'deep CCS'**,** can be achieved at a relatively low-cost premium compared to the benchmark of 90% capture efficiency [9]. Since deep CCS only necessitates a few modifications to traditional designs of 90% capture, it is a promising technology option to thoroughly decarbonize the fossil fuel energy industry [10]. Abated fossil fuels with deep CCS can play an important role in energy transition. Deep CCS also benefits other decarbonization pathways, such as electrification of transportation, buildings, and industries [1].

Few studies have conducted techno-economic assessments of deep CCS. Gao et al. (2019) performed Aspen Plus process simulation and showed that 99.1% may be achieved with a 20% increase in solvent flow rate and less than 5% increase in reboiler duty using the advanced flash stripper design, compared to 90% capture. Jiang et al. (2020) also proved that the advanced flash stripper layout can accomplish 99.7% emission reduction from coal combustion flue gas while increasing the cost of CO_2 avoided by $2.6 per metric ton of CO_2, compared to 90% capture. Du et al. (2021) used MEA solvent to achieve zero and negative CO_2 emissions from the flue gas of natural gas combined cycle (NGCC) and pulverized coal (PC) power plants, adopting simple and pump-around absorber intercooling configurations and increasing the height of the absorber. Deep CCS can achieve almost net-zero electricity emissions with minimal incremental cost compared to 90% CCS. This benefit may reduce the constraints on reaching the economy-wide net-zero emission target in the near future.

However, a more detailed analysis that integrates the process simulation and bottom-up engineering-economic model at the plant level is still lacking in the literature. Deep CCS is a promising technology that needs to be further evaluated at the power plant level. Therefore, this study configures an Aspen model of deep CCS and incorporated it into the Integrated Environmental Control Model (IECM) to provide another option for environmental control technology. A benchmark solvent, MEA, is used in a carbon capture system to remove 90–99% CO_2 from the flue gas of PC and NGCC power plants. The major performance parameters, including the flue gas flow rates, CO_2 concentration, lean loading, and flue gas temperature, are varied to analyze their effects on the overall plant performance and cost. Based on the key inputs and outputs of a wide range of simulation scenarios, a multivariable regression analysis is performed to develop a reduced-order performance model. Moreover, this study presents a bottom-up engineering-economic model in conjunction with the performance modeling results to evaluate the cost metrics, including capital costs, operating and maintenance costs, added levelized cost of electricity (LCOE), and the cost of CO_2 avoided. Finally, the integrated techno-economic model is incorporated into the IECM as the ‘Deep Carbon Capture’ module.

2. MEA-based CO_2 Capture System

The MEA-based CO_2 capture system consists of flue gas cooling, CO_2 absorption and stripping, and CO_2 compression units. This section discusses model development, process chemistry, process description, equipment, and model and base case assumptions and results in detail.

2.1. Model Development

Aspen Plus® is used to simulate the MEA-based carbon capture process from the flue gas of PC and NGCC power plants. Aspen provides rate-based and equilibrium-based model options for process simulations. Compared to the equilibrium-based model, the rate-based model is more rigorous and offers more accuracy [13]. Thus, in this study, the rate-based model is adopted to simulate the CO_2 capture process at various levels of removal efficiency. This model requires rigorous initialization, including column dimensions, packing specifications, absorbent properties, and liquid holdup. Rate-based distillation uses a mathematical model that includes mass transfer, energy transfer, phase equilibrium, material and energy balances, and summation equations to support its rate-based computations.

Bravo (1985) mass transfer correlation is used for the mass transfer coefficient and the interfacial area factor. The Bravo (1985) mass transfer correlation is tested for various experimental input parameters, and it showed close matches with the experimental CO_2 removal efficiency and absorber temperature profile at an interfacial area factor of unity. Similarly, the Chilton & Colburn (1934) heat transfer coefficients method predicted the experimental measurement accurately. For packed columns, random and structured packing materials can be used. In CO_2 capture processes with random packing materials, the reboiler duty is relatively higher than the structured ones [16] . Furthermore, the structured packing materials have low pressure drop, high contact efficiency, light weight, and the smallest possible column diameter. In structured packing materials, Mellapak materials are widely used in practice, and have significantly lower pressure drop compared to conventional packing materials [17]. Therefore, Mellapak structure packing material is utilized in the process. Table 1 shows the major model assumptions that remain consistent across all the scenarios.

Table 1: Major Model Assumptions for Carbon Capture System.

Model Category	Selected Model(s) for ASPEN Plus Process Simulation	Description
Absorber and Stripper	Rate-based Model	The model requires configuration of packing specifications, column dimensions, liquid holdup, and absorbent properties
Heat & Mass Transfer Correlations	Bravo (1985)	The model predicts the Mass Transfer Correlation and Interfacial Area Factor
	Chilton & Colburn (1934)	The method calculates the Heat Transfer Coefficient
Chemistry Model	Chemical Equilibrium and Kinetic	Chemical Equilibrium is assumed for all the ionic reactions except for those of CO_2 with OH^- and CO_2 with MEA, which are Kinetic
Vapor Properties	Redlich–Kwong Equation of State	The model calculates the Fugacity Coefficients
Liquid Properties	Electrolyte Non-Random Two Liquid (ENRTL)	The model calculates the Activity Coefficient, Gibbs free Energy, Enthalpy, and Entropy
Flow Model	Mixed Flow Model	In the Mixed Flow model, the bulk properties for each phase are assumed to be the same as the outlet conditions for that phase leaving that stage

An electrolyte non-random two-liquid (ENRTL) method is employed to calculate liquid properties, such as Gibbs free energy, activity coefficient, entropy, and enthalpy. As the CO_2 capture process chemistry includes the ionic components, the ENRTL method is suitable for such types of components. It defines the solution chemistry, input parameters for the equilibrium constants, and retrieves the binary interaction parameters. For the vapor properties, the Redlich-Kwong (RK) equation of state is adopted. The selected Henry components include O_2, N_2, CO_2, H_2S, and Argon (Ar). All the components used in this process are listed in Table 2.

Table 2: Components Used in the Process Modeling of Carbon Capture.

Component Name	Formula	Unit Operation
Carbon Dioxide	CO_2	All
Water	H_2O	All
Oxygen	O_2	All
Nitrogen	N_2	All
Argon	Ar	All
Bicarbonate Ion	HCO_3^-	All
Carbonate Ion	CO_3^{2-}	All
Hydroxide Ion	OH^-	All
Hydronium Ion	H_3O^+	All
Monoethanolamine (MEA)	C_2H_7NO	All
Monoethanolamine Ion (MEA^+)	$C_2H_8NO^+$	All
Methyl Carbamate ($MEACOO^-$)	$C_3H_6NO_3^-$	All

2.2. Process Chemistry

All the reactions occurring in the CO_2 capture process are taken from the Aspen Built-in example [18]. A chemistry model with an ID of "MEA" is used to model the electrolyte solution chemistry. In the chemistry model (MEA), chemical equilibrium is assumed for all the ionic reactions. Moreover, a reaction model with an ID called "MEA-REA" is also created. All the reactions in "MEA-REA" are presumed to be in equilibrium except those of CO_2 with MEA and CO_2 with OH^-. Tables 3 and 4 show the chemistry and reaction models.

Table 3: Reactions Occurring in the Absorber Column (ID: MEA).

Serial No.	Type	Reactions
1	Equilibrium	$2H_2O \leftrightarrow H_3O^+ + OH^-$
2	Equilibrium	$CO_2 + 2H_2O \leftrightarrow HCO_3^- + H_3O^+$
3	Equilibrium	$HCO_3^- + H_2O \leftrightarrow CO_3^{2-} + H_3O^+$
4	Equilibrium	$H_2O + MEAH^+ \leftrightarrow MEA + H_3O^+$
5	Equilibrium	$MEACOO^- + H_2O \leftrightarrow MEA + HCO_3^-$
6	Equilibrium	$H_2O + H_2S \leftrightarrow HS^- + H_3O^+$
7	Equilibrium	$H_2O + HS^- \leftrightarrow S^{2-} + H_3O^+$

Table 4: Reactions Occurring in the Stripper Column (ID: MEA-REA).

Serial No.	Type	Reactions
1	Equilibrium	$H_2O + MEAH^+ \leftrightarrow MEA + H_3O^+$
2	Equilibrium	$2H_2O \leftrightarrow H_3O^+ + OH^-$
3	Equilibrium	$HCO_3^- + H_2O \leftrightarrow CO_3^{2-} + H_3O^+$
4	Equilibrium	$H_2O + H_2S \leftrightarrow HS^- + H_3O^+$
5	Equilibrium	$H_2O + HS^- \leftrightarrow S^{2-} + H_3O^+$
6	Kinetic	$CO_2 + OH^- \leftrightarrow HCO_3^-$
7	Kinetic	$HCO_3^- \leftrightarrow CO_2 + OH^-$
8	Kinetic	$MEA + CO_2 + H_2O \leftrightarrow MEACOO^- + H_3O^+$
9	Kinetic	$MEACOO^- + H_3O^+ \leftrightarrow MEA + CO_2 + H_2O$

In the built-in model, Gibbs free energy is used to calculate the equilibrium constants for reactions 1 to 5 (chemistry model). For No. 6-7 reactions reported in Table 4, equilibrium constants are taken from the work of Austgen et al. (1989). From the works of Pinsent et al. (1956), the kinetic parameters are obtained for the reactions 6-9. For kinetic reactions 6-9 (MEA-REA) calculation, Equation 1 is used:

$$r = k\left(\frac{T}{T_0}\right)^n \cdot \exp\left[\left(\frac{-E}{R}\right) \cdot \left(\frac{1}{T} - \frac{1}{T_0}\right)\right] \cdot \prod_{i=1}^{N} C_i^{ai} \quad (1)$$

where,

r = Rate of reaction (kmole/m^3-s)

E = Activation energy (cal/mole)

T = Absolute temperature (Kelvin)

N = Number of components

k = Pre-exponential factor; n is the temperature exponent

R = Gas constant (J/mole-k)

C_i = Component concentration (molarity, kmole/m^3, mole fraction)

T_0 = Reference temperature (Kelvin)

2.3. Process Description

Figure 1 shows a flowsheet diagram of CO_2 capture per train, processing half of the flue gas of a 650 MW_g coal-fired power plant. It consists of 2 absorbers, a heat exchanger, a cooler, makeup mixers, and a stripper. The overall CO_2 capture system of the 650 MW_g coal-fired power plant is composed of two parallel process trains.

The flue gas needs to be denitrified, desulphurized, and cooled before being flowed into the bottom of the absorber. It is assumed that all the necessary operations on flue gas are performed before entering the CO_2 capture process. Two absorbers, ABS-1 and ABS-2, are designed to treat half of the flue gas from a 650 MW_g coal-fired power plant. The flue gas is split into FG-1 and FG-2, as shown in Figure 1, with FG-1 flowing to ABS-1 and FG-2 flowing to ABS-2. In addition to the flue gas, an aqueous amine solution is directed at the top of the absorbers (LIN-1 to ABS-1 and LIN-2 to ABS-2). This solution flows downward in the absorber columns to capture CO_2 from the flue gas. After absorbing CO_2, the lean streams exit the bottoms of the absorbers as carbon-rich streams (ROUT-1 and ROUT-2), while the clean CO_2-flue gas streams, CG-1 and CG-2, are emitted from the tops of the absorbers. The CG-1 and CG-2 streams are mixed and pumped into the wash column to prevent solvent emissions. Similarly, both the carbon-rich solution streams are mixed and pass through the heat exchanger (HX). The mixed stream is then introduced into the stripper column as RIN. Here, the RIN stream is heated using a reboiler to release the CO_2 gas. The CO_2 gas is cooled in the condenser to release water vapor and then send to the compression unit. The remaining solution is drained from the bottom of the stripper as a hot lean-out stream called BOTL-1. This stream exchanges heat in HX with the other stream. The logarithmic mean temperature difference (LMTD) of 10 °C is considered for the HX. After adding MEA and water to the lean-out stream, it passes through the cooler. The lean-out solution is split into two equal streams (LIN-1 and LIN-2). Both streams are delivered back to the absorbers to continue the cycle.

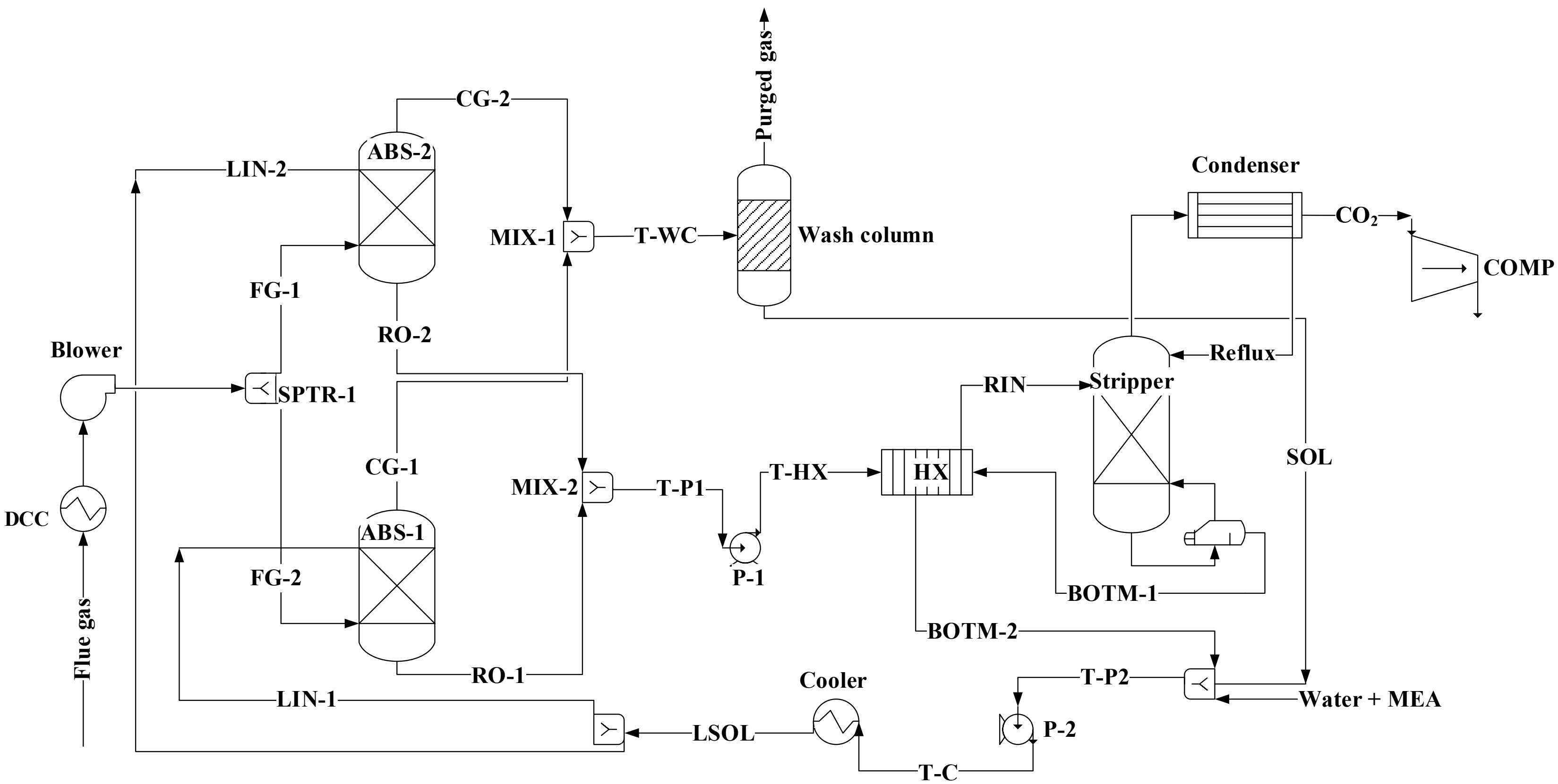


Figure 1: *Flowsheet diagram of one process train of CO_2 capture and compression from a 650 MW_g coal-fired power plant.*

2.4. Process Equipment

The major equipment in CO_2 capture and compression processes are the following:

Direct contact cooler (DCC): The temperature of flue gases exiting from the power plant is quite high and undesirable for the CO_2 absorption process. The exothermic reactions occurring in the absorber column are favored by the low flue gas temperature. For maximum solvent loading, the desirable temperature and pressure of flue gas for amine-based CO_2 capture absorption is 40 °C and 1 atm total pressure [21]. Therefore, a DCC is used to reduce the flue gas temperature. In gas-fired and many coal-fired power plants that lack wet scrubbers for the removal of SO_2, a DCC must be installed to lower the temperature of the flue gas stream to an acceptable level. When a wet flue gas desulfurization (FGD) unit is installed upstream of the amine system in a coal-fired power station, the wet scrubber significantly reduces the temperature of the flue gas, possibly negating the need for an additional cooler.

Flue gas blower: A pressure drop occurs to the flue gas as it flows upward in the absorber columns. Flue gas blowers help to overcome the pressure drop of the flue gas in it.

Splitters: Splitter 1 is used to split the flue gas into two equal streams: FG-2 and FG-2. Similarly, Splitter 2 is installed to split the lean solution into two equal streams (LIN-1 and LIN-2).

Absorber column: In this distillation column, the aqueous MEA solution encounters flue gas to absorb CO_2 from it. The column is either plate-type or a packed bed. Although plate-type absorbers have been successfully operated, packed columns tend to allow for decreased pressure drop, higher gas throughput, improved gas contact efficiency, and less potential for foaming. A packed column is used with Mellapak structured packing material to provide better results [16]. The absorber column is operated at a pressure of 1 bar [22].

Due to the high volume of flue gas from a coal-fired power plant, multiple absorbers and stripper columns are used. Installing more than one absorber column can improve the turndown ratio of the CO_2 capture process. Since most power plants operate under flexible load conditions, integrating multiple absorber columns into the carbon capture unit allows the system to adjust more efficiently to varying power outputs, maintaining optimal capture performance and energy efficiency. Based on the flow rates of flue gas for various power plants, the number of absorbers varies from 2 to 5 to accommodate the flue gas. It is essential to prevent flooding in the absorber and stripper columns to ensure proper downward liquid flow, maintain efficient mass transfer, and minimize pressure drop. Thus, both the columns are designed under the constraint of a flooding approach of no more than 70% in the column [23]. The absorber and stripper diameters are kept below the maximum diameter of 12.2 m [24], and the column's height-to-diameter ratio is kept higher than 1 [25]. A column with a smaller diameter is easier to transport, handle during installation, and manufacture [26]. A RadFrac rate-based model is used for the absorption operation, which performs rigorous calculations for the rating and design. The RadFrac model is used for absorption, stripping, ordinary distillation, extractive and azeotropic distillation, reactive distillation, and three-phase distillation operations. It can model the columns in which chemical reactions occur. Reactions might be fixed for conversions, equilibrium, rate-controlled, or electrolytic.

Absorber Intercooler: Absorber intercoolers helped improve the absorption rate and reduce the solution temperature in the column [8]. They collect the solution from the upper section, cool it, and send it to the lower section. Each absorber is equipped with one intercooler, except for the one used in the 99% CO_2

capture cases, which has two intercoolers. This design is necessary because the 99% CO_2 capture cases require a higher cooling duty to remove a greater amount of CO_2. Absorber intercoolers are used in the bottom of the column to provide better results in terms of CO_2 capture rate. In CO_2 capture from PC power plants, absorber intercoolers are typically installed at Stage 27 for 90%, 95%, and 97% capture rates, while for a 99% capture rate, they are positioned at Stages 13 and 26. In NGCC power plants, absorber intercoolers are most commonly located at Stage 54 of the absorber.

Mixers: Mixer 1 is used to combine the treated flue gas streams: CG-1 and CG-2. Mixer 2 is applied to combine the carbon-rich streams: RO-1 and RO-2. Mixer 3 combines the lean stream (BOTM-1), liquid stream from the wash column, and the make-up sorbent.

Wash column: The treated flue gases (CG-1 and CG-2) exiting from the absorber top contain vaporized or entrained MEA. In the U.S., the National Institute for Occupational Safety and Health (NIOSH) has established a permissible exposure limit of 3 ppmv for MEA to help protect workers' health and safety [27]. To prevent MEA emissions above 3 ppmv, a wash column is used. It is a flash vessel that separates liquid and gas streams. The liquid stream (SOL) flowed to Mixer 3 and mixed with the make-up sorbent while the gas stream is released to the atmosphere.

Rich solution pump: The CO_2 rich solution pump (P1) with 75% efficiency is used to elevate the pressure of the solution (RO-1 and RO-2) exiting from the bottom of the absorber to eliminate the tendency of acid gas breakout in the lean/rich heat exchanger and meet the stripper's operating pressure and height requirements. In this study, pump P1 is used to increase the pressure of the carbon-rich solution up to 2 bar to match the operating pressure of the stripper column, which is 1.9 bar.

Lean/rich heat exchanger: The carbon-rich solution needs to be preheated before entering as RIN to the stripper column to help in reducing the reboiler duty. On the other hand, the CO_2 lean solution coming out of the bottom of the stripper needs to be pre-cooled before it goes to the lean solution cooler. For this purpose, a cross-heat exchanger (HX) is used where these two sorbent streams are passed. The LMTD of 10 °C is considered for the HX. The reboiler duty can be reduced if a lower (5 °C) approach for the LMTD is specified. However, this can increase the required surface area of HX.

Stripper/ Regenerator column: This vessel is used to remove CO_2 from the carbon-rich solution using steam. The heat supply reverses the absorption reactions and breaks down the carbamate compound formed between MEA and dissolved CO_2. It is a packed column like the absorber where the carbon-rich solution flows down, and steam rose to strip CO_2 from the solution. The reboiler temperature is kept below 122 °C to avoid solvent thermal degradation [28]. The stripper is fixed at a pressure of 1.9 bar, resulting in the reboiler temperature below 122 °C. A design specification option is used to fix the condenser temperature at 30 °C, while varied the reflux ratio. Both stripper and absorber columns diameter are designed in such a way to manage below 70% flood.

Reboiler: A kettle-type reboiler is used to heat the CO_2 loaded solvent by utilizing the power plant's low-pressure steam. It is attached to the bottom of the stripper, where the lean amine solution flows by gravity and vaporizes a portion of it. The vapor is provided back to the stripper to regenerate the carbon-rich sorbent. The hot lean solution coming out from the reboiler to the HX.

Reflux condenser: After heating the carbon-rich stream to strip out CO_2, the steam that is left at the top of the stripper contains mainly CO_2 gas and water vapors. It is passed through the condenser to condense the water vapors and purify the CO_2 product gas. The condensed stream is refluxed back to the stripper, and the CO_2 product gas is directed to the compressor unit. Cooling water at 21 °C is used to reduce the temperature of the CO_2 product stream up to 30 °C. The lower the condenser temperature, the higher the purity of CO_2.

Lean solution cooler: The lean solution temperature is higher even after passing through the HX; it must be further cooled down to reach the acceptable level. Cooling water at 21 °C is used to reduce the temperature of lean solution to 40 °C in the cooler.

Lean solution pump: After flowing through the HX, Mixer 3, and cooler, the lean solution experiences a drop in pressure. To overcome it, a lean solution pump (P2) is used to increase the pressure of the lean solution (LIN-1 and LIN-2).

Compression unit: The main purpose of the compression unit is to compress the CO_2 product gas to an elevated pressure for transporting it over a long distance via pipelines. This unit consists of a multi-stage compressor, interstage coolers, flash drums, and an aftercooler. The discharge pressure and temperature of product CO_2 from the capture unit are 152.7 bar and 30 °C, respectively, for all the cases. Thus, the outlet stage pressures, compression specifications, and stage pressure ratios are identical for each case with the key differentiator being the flow rate of inlet CO_2.

The design configuration is taken from the NETL report entitled "Cost and Performance baseline for fossil energy plants: Volume 1: Bituminous coal and natural gas to electricity" [29]. It consists of an eight-stage centrifugal compressor with a feed stream at stage 1. Table 5 shows compressor efficiency, stage discharge pressure, aftercooler temperature, and CO_2 purity.

Table 5: Parameters and Assumptions for Compression Unit.

Parameter	Value			
Compressor Efficiency	80%			
Compressor Stages	8			
Aftercooler Temperature (°C)	30			
Outlet Pressure of Each Stage (bar)	1st 2.6	2nd 5.9	3rd 12.2	4th 25.1
	5th 39.7	6th 63.4	7th 98.7	8th 152.7
CO_2 Purity (mole%)	99.8			

2.5. Base Case Assumptions of Coal-fired Power Plants

To evaluate different scenarios, base case assumptions are made to explore variations in the flue gas flow rate, flue gas temperature, CO_2 lean loading, CO_2 concentration in the flue gas, etc. Table 6 shows a summary of all the equipment data for the base case.

Table 6: Base Case Assumptions of CO_2 Capture from a 650 MW_g Coal-fired Power Plant.

Parameter	Value				
Inlet Flue Gas					
Flow Rate (tonne/hr)	3,305.8				
Composition (mole%)	CO_2	H_2O	O_2	N_2	Ar
	12	15	4.2	68	0.8
Temperature (°C)	45				
Pressure (bar)	1.10				
Number of CO_2 Capture Process Trains	2				
CO_2 Capture Rate (%)	90, 95, 97, and 99				
Heat Exchanger					
Exchanger LMTD (°C)	10				
Lean Solution Stream					
MEA Concentration (wt.%)	30				
Lean Loading (mole CO_2/mole MEA)	0.20				
Temperature (°C)	40				
Pressure (bar)	1.10				
Solvent Circulation Pumps Efficiency (%)	75%				
Wash Column					
MEA Emission from Wash Column (ppm)	< 3				
Inlet/Outlet Cooling Water Temperatures (°C)	21.11/ 32.22				
Absorber Column Specification					
Pressure (bar)	1				
Stages	30				
Packing Material	MellaPak, Standard, 250Y				
Base Flood	70%				
Number of Intercooler (per absorber)	1 (Except 99% Case*)				
Stripper Column Specification					
Pressure (bar)	1.90				
Number of Stages	40				
Packing Material	MellaPak, Standard, 750Y				
Condenser Temperature (°C)	30				
Inlet/Outlet Cooling Water Temperatures (°C)	21.11/ 32.22				
Stripper Reboiler temperature (°C)	< 122				
Cooler for Lean Solution					
Temperature (°C)	40				
Inlet/Outlet Cooling Water Temperatures (°C)	21.11/ 32.22				
CO_2 Compression Specification					
Isentropic Efficiency (%)	80%				
Number of Stages	8				
CO_2 Product Gas Purity (mole%)	99.80				
CO_2 Product Gas Pressure (MPa)	15.25				

2.6. Results and Discussion of Coal-fired Power Plants

The flue gas from a 650 MW_g coal-fired power plant is considered, as shown in Table 7. The flue gas flow rate is 3305.81 tonne/hr (115,203.53 kmole/hr) and contain a CO_2 concentration of 12 mole%. To treat that high volume of flue gas, 4 absorbers and 2 stripper columns are used.

Table 7 shows the input and output summary of major parameters and equipment. The lean solution flow rate and absorbers packing volume are varied to achieve the desired capture rates of 90, 95, 97, and 99%. The energetic optimum for lean loading is 0.20 mol CO_2/mol MEA [30]; therefore, the CO_2 lean loading is fixed at 0.20 mol CO_2/mol MEA in this study. The temperature and pressure of lean solution are designed to be 40 °C and 1.1 bar, respectively. The L/G ratio, reboiler duty, cooling water requirement, pumps and compressor power requirement, and the design dimensions of absorbers and stripper columns increase with the CO_2 capture rate.

Table 7: Summary of Major Performance Parameters of CO_2 Capture Processes.

Parameter	**Capture Rate (%)**			
	90	**95**	**97**	**99**
Lean Solution Flow Rate (kmole/s)	100.43	107.73	112.0	117.67
Liquid-to-Gas Ratio (L/G) (mole/mole)	3.14	3.37	3.50	3.68
Absorber Column (m) (each)[a]	D=11.7, H=11.8	D=11.9, H=13	D=12, H=14	D=11.8, H=16
Absorber Pressure (bar)	1.0	1.0	1.0	1.0
Stripper Column (m) (each)	D=11.7, H=12	D=11.9, H=12.5	D=12, H=13	D=12.2, H=13
Stripper Pressure (bar)	1.9	1.9	1.9	1.9
Condenser Temperature (°C)	30	30	30	30
Pumps Power (MWe)	1.0	1.29	1.35	1.44
Reboiler Duty (MW_{th})	525.07	560.68	579.93	608.40
Number of Intercoolers	4	4	4	8
Absorber Intercooler Cooling Water Requirement (tonne/hr)	7,760	7,760	7,760	15,520
Wash Column Cooling Water Requirement (tonne/hr)	8,520	9,621	9,721	3,543
Lean solution Cooler Cooling Water Requirement (tonne/hr)	7,128	9,495	11,598	15,174
Condenser Cooling Water Requirement (tonne/hr)	11,116	11,969	12,490	13,370
CO_2 Capture Unit Cooling Water Requirement (tonne/hr)	34,520	38,845	41,569	47,607
Compression Cooling Water Requirement (tonne/hr)	5,263	5,556	5,673	5,790
Total Cooling Water Requirement (tonne/hr)	39,783	44,401	47,242	53,397

MEA Make-up (kg/hr)	12.42	12.37	12.11	11.56
Compressor Power (MWe)	48.66	51.36	52.44	53.52
Amount of CO_2 Captured (tonne/hr)	546.69	577.07	589.72	601.35

a. D = diameter; H = packing height.

From the performance results, key performance metrics are identified, including flue gas flow rate and temperature, CO_2 concentration, CO_2 lean loading, and L/G ratio. A sensitivity analysis is needed to further evaluate the effect of these performance metrics on the capture process.

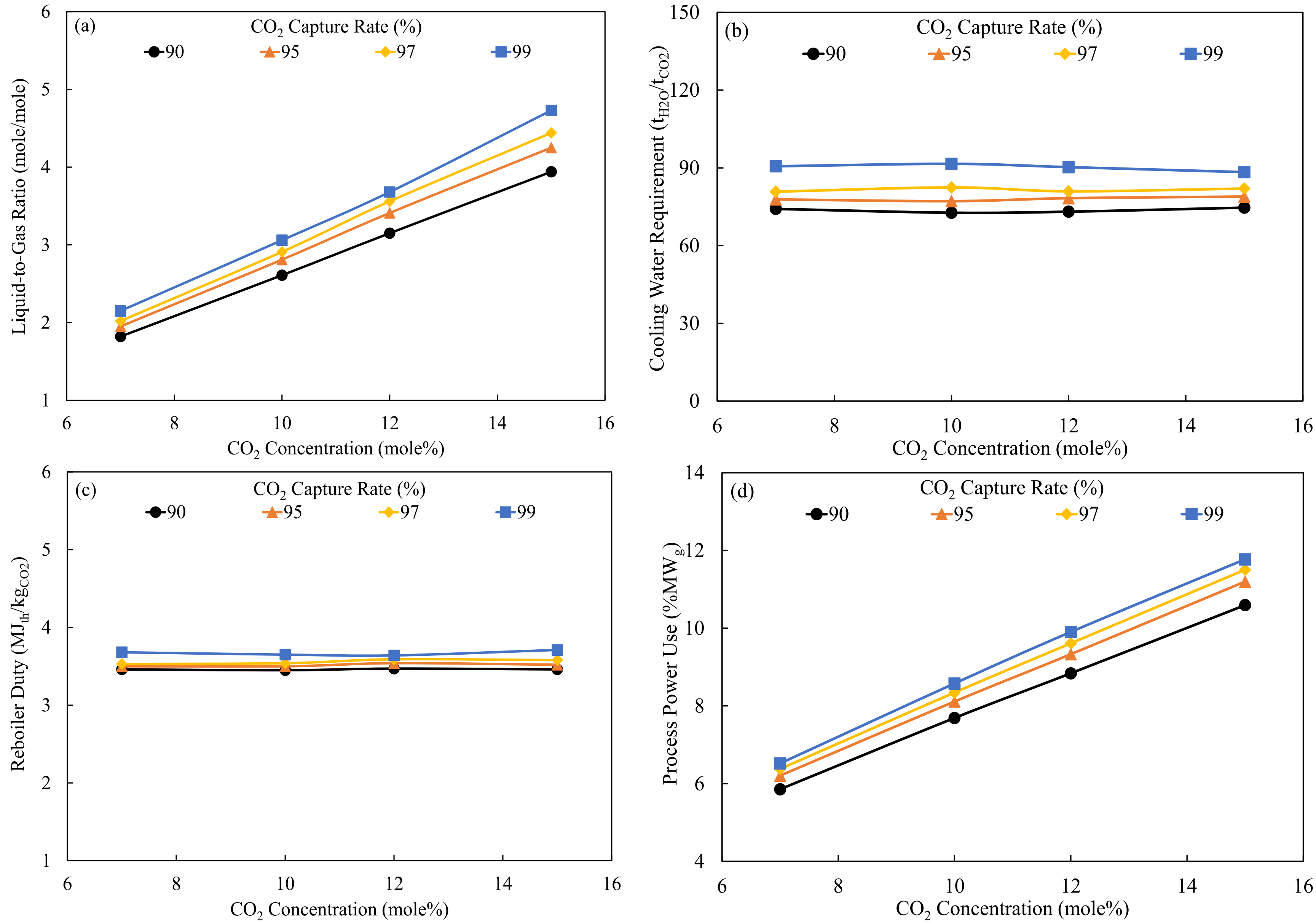
2.6.1 Sensitivity Analysis for Coal-fired Power Plants

This section is dedicated to sensitivity analysis by varying the flue gas flow rate, flue gas temperature, CO_2 lean loading, and CO_2 concentration in the flue gas to uncover their impact on the overall performance of the capture system.

2.6.1.1 Effects of CO_2 Concentration

The concentration of CO_2 in the flue gas has a significant effect on the CO_2 capture process. To analyze the major performance parameters, the CO_2 concentrations in flue gas are varied from 7 mole% to 15 mole%. The major performance parameters affected by the CO_2 concentrations include the L/G ratio, cooling water requirement, reboiler duty, process power use, and the volumes of absorption and stripping packing. Figure 2(a) shows that the L/G ratio increases with the CO_2 concentration and capture rates. This is because at a higher capture rate, a higher solution flow rate is required while the flue gas flow rate remains the same. In the 90% CO_2 removal rate, the L/G ratio increases from 1.82 to 3.94 mole/mole by increasing the CO_2 concentration from 7 mole% to 15 mole% in the flue gas. Similarly, it increases from 1.95 to 4.25 mole/mole, 2.02 to 4.44 mole/mole, and 2.15 to 4.73 mole/mole in 95, 97, and 99% CO_2 removal rates, respectively.

The cooling water requirements only increase with the capture rates. It is due to the cooling down of the high volume of treated gas in the wash column, the high solution flow rate in the cooler, and the elevated flow rate of CO_2 in the stripper's condenser and compressors' intercoolers. The cooling water requirements are almost 73, 78, 81, and 90 tonne H_2O/tonne CO_2 captured in 90, 95, 97, and 99% CO_2 capture processes, respectively. A 99% capture rate requires 2 intercoolers in each absorber; therefore, its cooling duty increases exponentially as shown in Figure 2(b).



(a)
CO2 Capture Rate (%)
90
95
97
99
Liquid-to-Gas Ratio (mole/mole)
CO2 Concentration (mole%)
(b)
CO2 Capture Rate (%)
90
95
97
99
Cooling Water Requirement (tH2O/tCO2)
CO2 Concentration (mole%)
(c)
CO2 Capture Rate (%)
90
95
97
99
Reboiler Duty (MJth/kgCO2)
CO2 Concentration (mole%)
(d)
CO2 Capture Rate (%)
90
95
97
99
Process Power Use (%MWg)
CO2 Concentration (mole%)

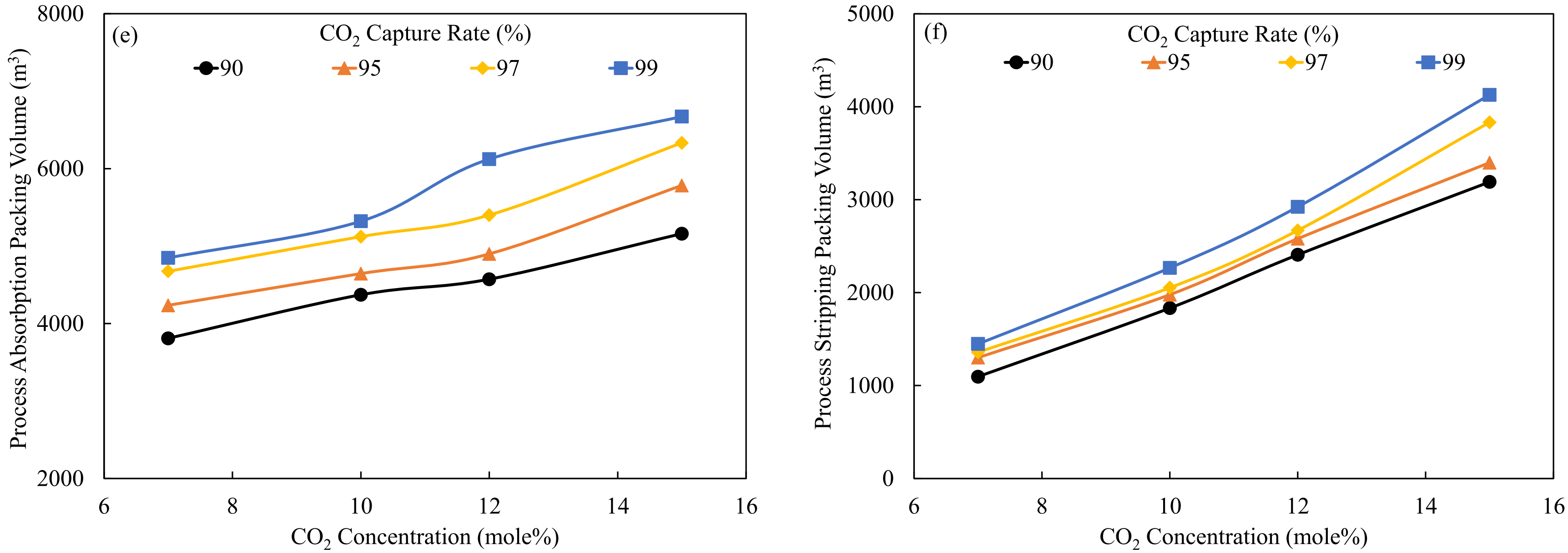


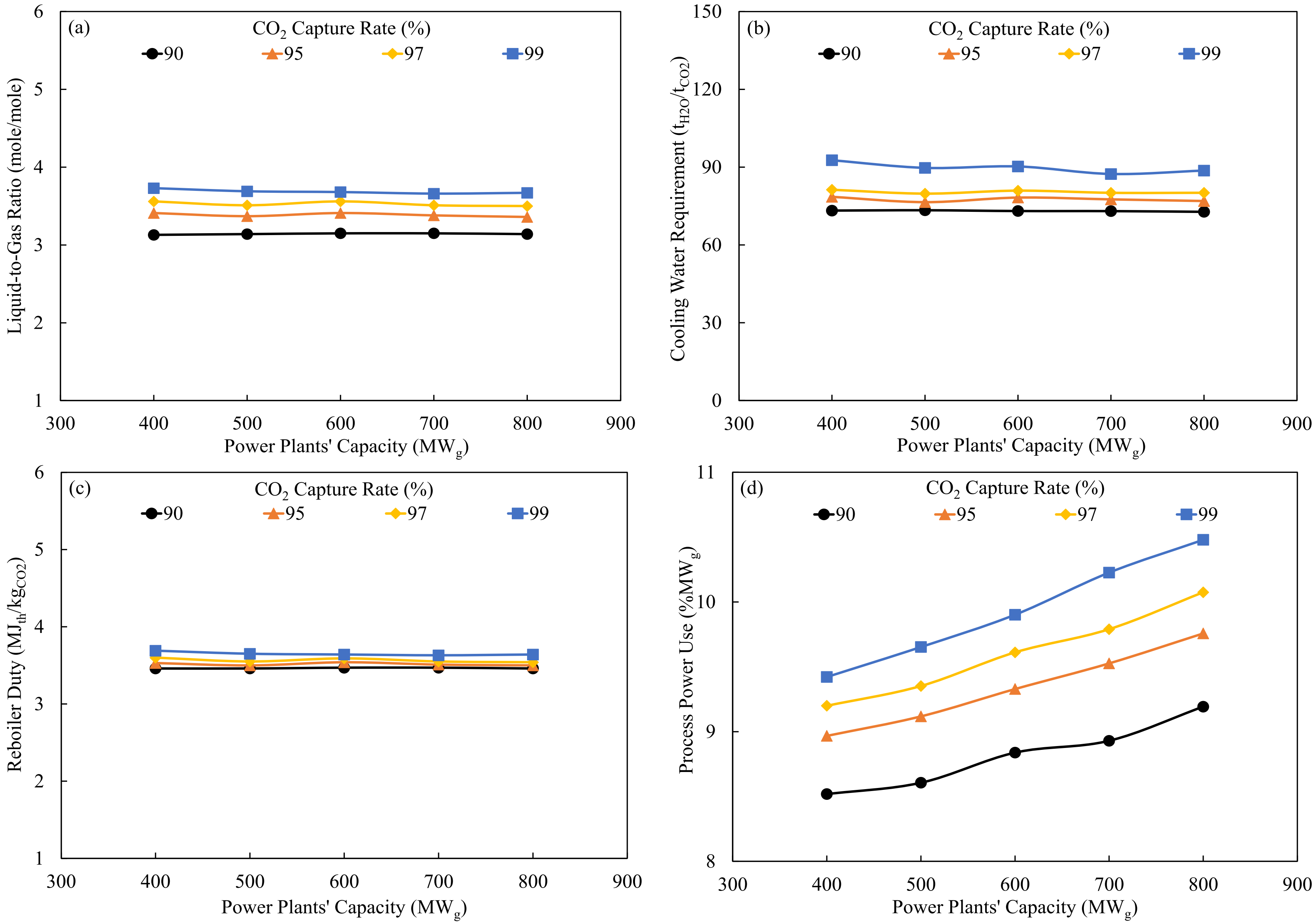
***Figure 2**: Effects of CO_2 concentrations on key performance parameters of coal power plants: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volumes, and (f) process stripping packing volumes.*

The normalized reboiler duty is almost the same for all CO_2 concentrations, having the same capture rates while increases only with the CO_2 capture rates as depicted in Figure 2(c). The 90, 95, 97, and 99% CO_2 capture rates require reboiler duties of 3.46, 3.51, 3.56, and 3.67 MJ_{th}/kg CO_2, respectively. The process power requirement is the sum of the power used by the blower, pumps, and compressors. Figure 2(d) illustrates that the power requirement increases with CO_2 concentrations and capture rates. In the higher CO_2 capture rates, the solution flow rate is higher, and hence the solution pumping rises. Additionally, a high capture rate means more CO_2 passes through the compressors, thus increasing the overall power requirement. In the 90% CO_2 removal rate, the parasitic load increases from 5.86 to 10.6 %MW_g by increasing the CO_2 concentration from 7 mole% to 15 mole%. In the same way, it increases from 6.2 to 11.2 %MW_g, 6.38 to 11.51 %MW_g, and 6.52 to 11.77 %MW_g in the 95, 97, and 99% CO_2 capture rates, respectively.

Figure 2(e, f) shows that the process absorption and stripping packing volumes increase with the CO_2 concentration and capture rates to accommodate higher CO_2 concentrations and solution flow rates. The process absorption packing volumes increases from 3810 to 5161 m^3 with CO_2 concentration at the 90% CO_2 removal rate. Likewise, in the 95, 97, and 99% removal rates, it increases from 4236 to 5784 m^3, 4676 to 6334 m^3, and 4849 to 6673 m^3, respectively. The process stripping packing volume increases from 1095 to 3191 m^3, 1301 to 3398 m^3, 1359 to 3832 m^3, and 1448 to 4128 m^3 in the 90, 95, 97, and 99% CO_2 removal rates, respectively. In both the 12 mole% and 15 mole% CO_2 concentration cases, 5 absorbers and 3 strippers are required. While 7 mole% and 10 mole% required fewer than 5 absorbers and 3 strippers. Therefore, the packing volumes of both 12 mole% and 15 mole% are higher.

2.6.1.2 Effects of Flue Gas Flow Rate

In this subsection, various coal-fired power plants ranging from 400 MW_g to 800 MW_g are considered to analyze the L/G ratio, reboiler duties, cooling water requirements, amine scrubber power use, and process absorption and stripping packing volumes. The inlet CO_2 concentration is fixed at 12 mole%. Figure 3(a) shows that the L/G ratio only increases with the capture rates because a higher solution flow rate is needed for a higher capture rate, while the flue gas flow rate remains the same. The L/G ratio is the same for different power plants having the same capture rates because when the solution flow rate increases, the flue gas flow rate also increases. It is almost 3.14 mole/mole in 90%, 3.38 mole/mole in 95%, 3.51 mole/mole in 97%, and 3.68 mole/mole in 99% CO_2 removal rates.


(a)
CO2 Capture Rate (%)
90
95
97
99
Liquid-to-Gas Ratio (mole/mole)
Power Plants' Capacity (MWg)
(b)
CO2 Capture Rate (%)
Cooling Water Requirement (tH2O/tCO2)
Power Plants' Capacity (MWg)
(c)
CO2 Capture Rate (%)
Reboiler Duty (MJth/kgCO2)
Power Plants' Capacity (MWg)
(d)
CO2 Capture Rate (%)
Process Power Use (%MWg)
Power Plants' Capacity (MWg)

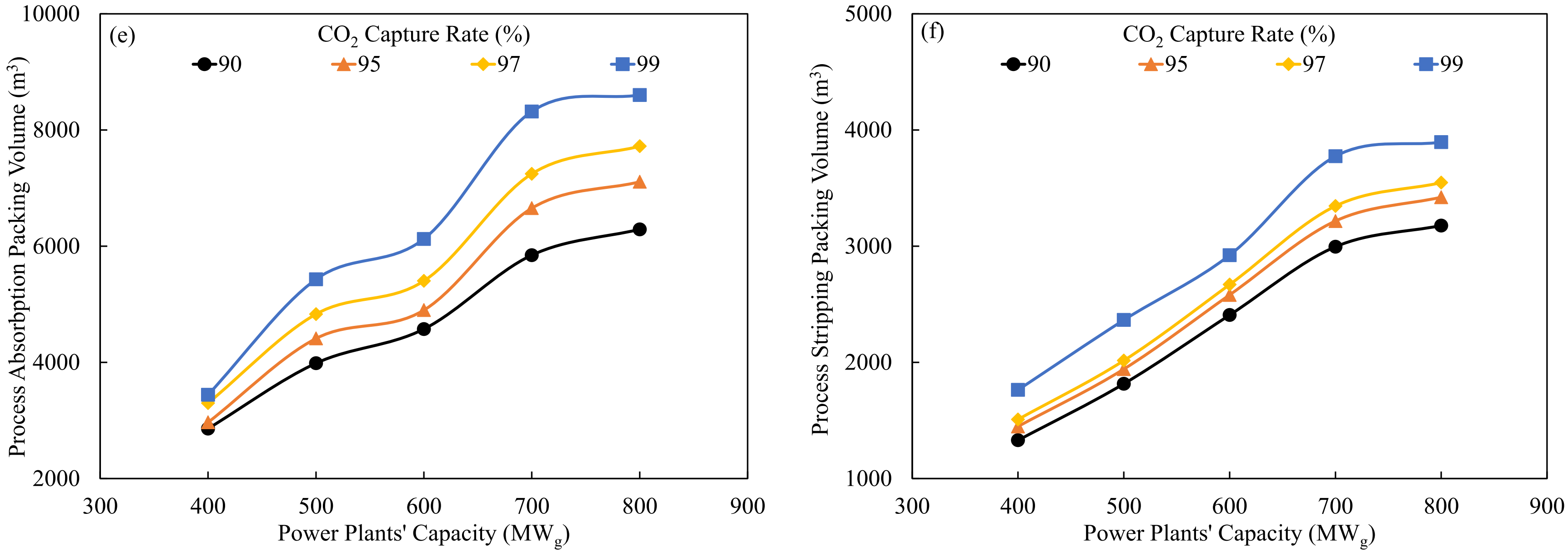


Figure 3: *Effects of power plant capacity on key performance parameters of coal power plants: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*

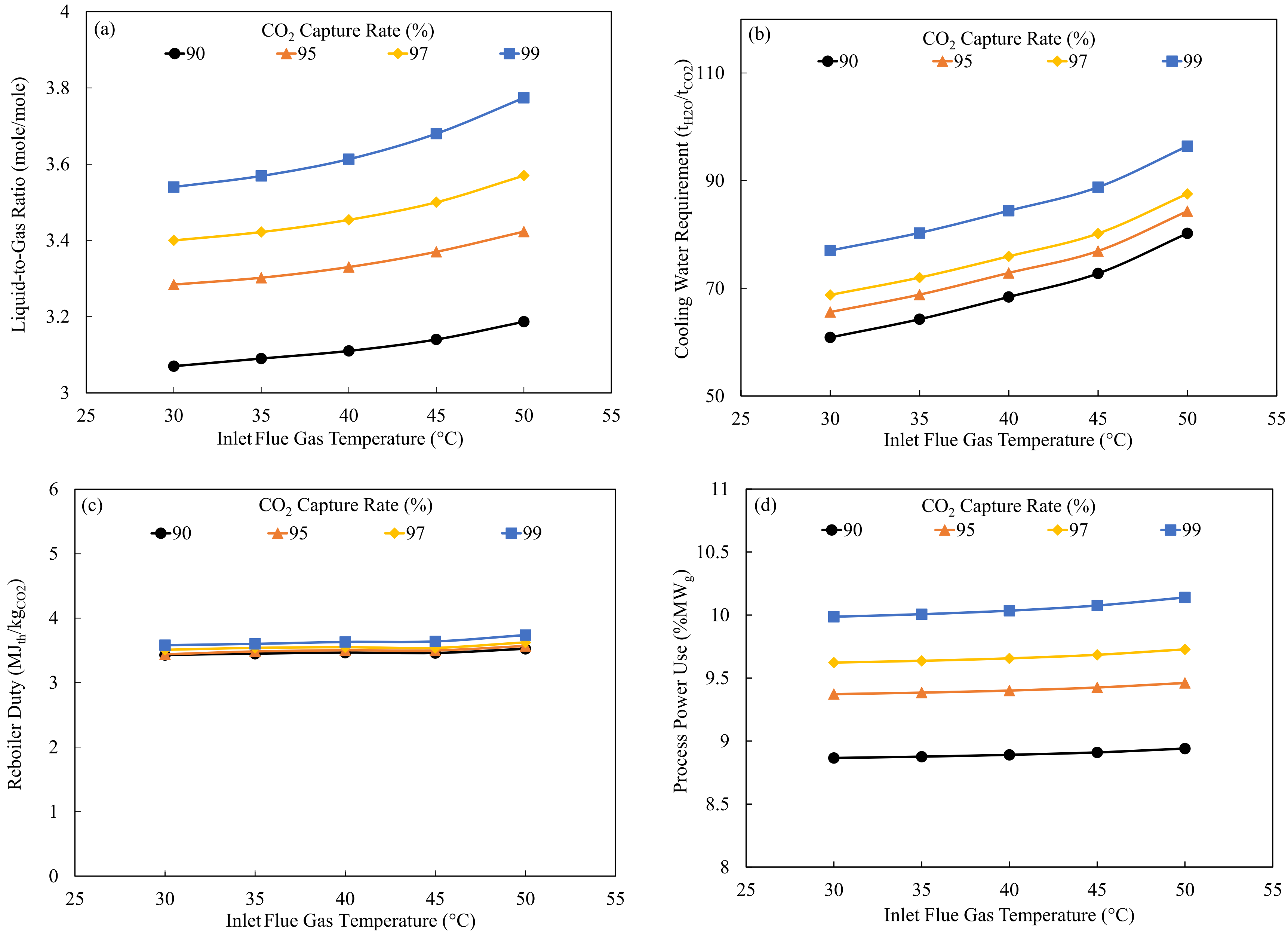
The normalized cooling water requirements have the same pattern as the L/G ratio. It only increases with the capture rates due to the cooling down of high volumes of the treated gas in the wash column, the high solution flow rate in the cooler, and the elevated flow rate of CO_2 in the stripper's condenser and compressors' intercoolers. The CO_2 removal processes of 90, 95, 97, and 99% require cooling water of 73, 78, 81, and 90 tonne H_2O/tonne CO_2, respectively. The reboiler duty is the same for all the power plants having the same capture rates, as shown in Figure 3(c). It only increases when the capture rates increase to remove a higher amount of CO_2 gas in the stripper. The reboiler duties required for the respective cases are explained in the previous section.

Power use increases with the power plant's gross output and capture rates. With increasing the power plant's capacity, the flue gas flow rates increase, and hence the blower and solution pumping require more power as depicted in Figure 3(d). Moreover, a high capture rate means more CO_2 passes through the compressors, thus its consumption also increases, and subsequently the overall power requirement rises. In the 90% CO_2 removal rate, it increases from 8.52 to 9.19 %MW_g by raising the power plants' capacity from 400 MW_g to 800 MW_g. Likewise, it increases from 8.97 to 9.76 %MW_g, 9.20 to 10.10 %MW_g, and 9.42 to 10.48 %MW_g in the 95, 97, and 99% CO_2 capture rates, respectively.

The process absorption and stripping packing volumes increase with the power plant's capacity and capture rates to accommodate the higher flue gas and solution flow rates. Figures 3(e) and 3(f) show that the process absorption and stripping packing volumes increase with the flue gas flow and capture rates to accommodate the higher CO_2 concentrations and solution flow rates. The process absorption packing volume increases from 2859 to 6289 m^3 with the increase of power plant's capacity from 400 MW_g to 800 MW_g in the 90% CO_2 removal rate. Likewise, in the 95, 97, and 99% removal rates, it increases from 2967 to 7108 m^3, 3298 to 7720 m^3, and 3441 to 8601 m^3, respectively. The process stripping packing volume increases from 1330 to 3177 m^3, 1448 to 3421 m^3, 1509 to 3547 m^3, and 1763 to 3895 m^3 in the 90, 95, 97, and 99% CO_2 removal rates, respectively. The absorption packing volume of the 400 MW_g case is lower as it requires 3 absorbers, while above 400 MW_g scenarios require more absorbers.

2.6.1.3 Effects of Flue Gas Temperature

To evaluate the major parameters, the inlet flue gas temperature varies from 30 °C to 50 °C while keeping the inlet CO_2 concentration at 12 mole% and the CO_2 lean loading at 0.20 mole/mole. Figure 4(a) indicates that the L/G ratio increases with the flue gas temperatures and capture rates. This is because of a higher flue gas temperature and capture rate, a higher solution flow rate is needed to capture the desired amount of CO_2. The normalized cooling water requirements have the same pattern as the L/G ratio, as shown in Figure 4(b). A 99% capture rate requires 2 intercoolers in each absorber; therefore, its cooling water requirement is higher compared to other capture rates. When CO_2 capture rates increase from 90% to 99%, the L/G ratio increases by 14% and the cooling water requirement increases by 19% on average for the given range of inlet flue gas temperatures. Increasing the inlet flue gas temperature from 30 to 50 °C increases the L/G ratio by 4.7% and the cooling water requirement by 22% on average for all the capture rates.



(a)
CO2 Capture Rate (%)
90
95
97
99
Liquid-to-Gas Ratio (mole/mole)
Inlet Flue Gas Temperature (°C)
(b)
CO2 Capture Rate (%)
90
95
97
99
Cooling Water Requirement (tH2O/tCO2)
Inlet Flue Gas Temperature (°C)
(c)
CO2 Capture Rate (%)
90
95
97
99
Reboiler Duty (MJth/kgCO2)
Inlet Flue Gas Temperature (°C)
(d)
CO2 Capture Rate (%)
90
95
97
99
Process Power Use (%MWg)
Inlet Flue Gas Temperature (°C)

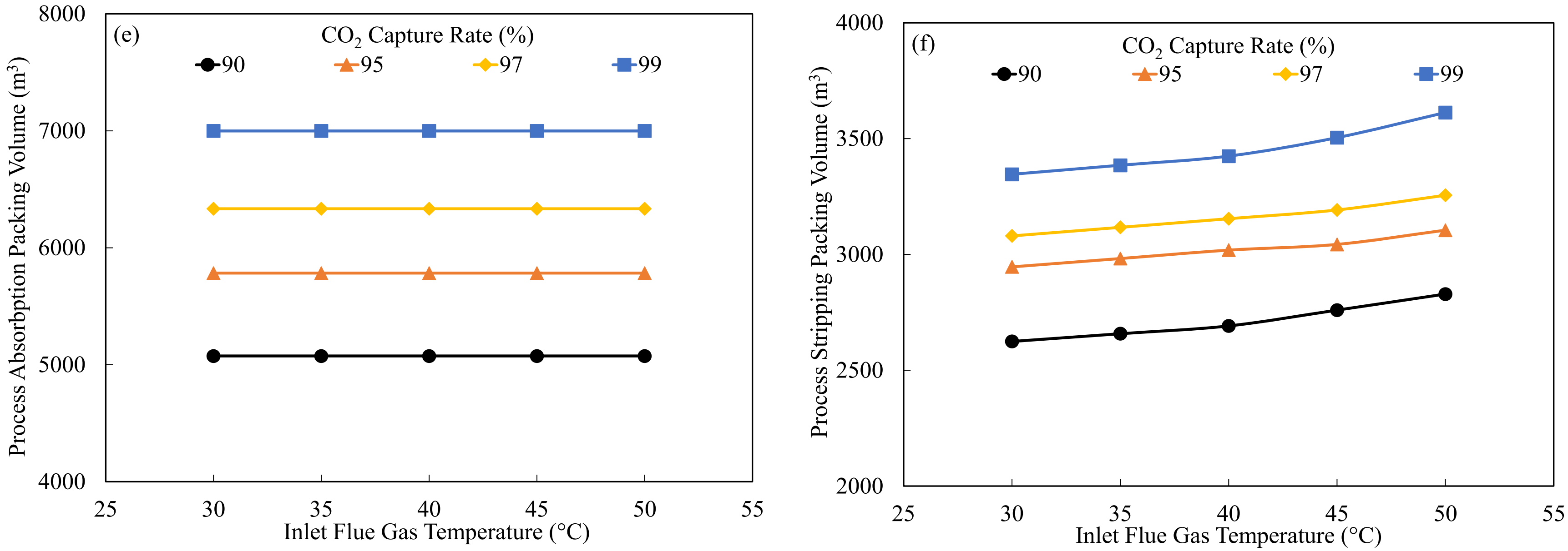


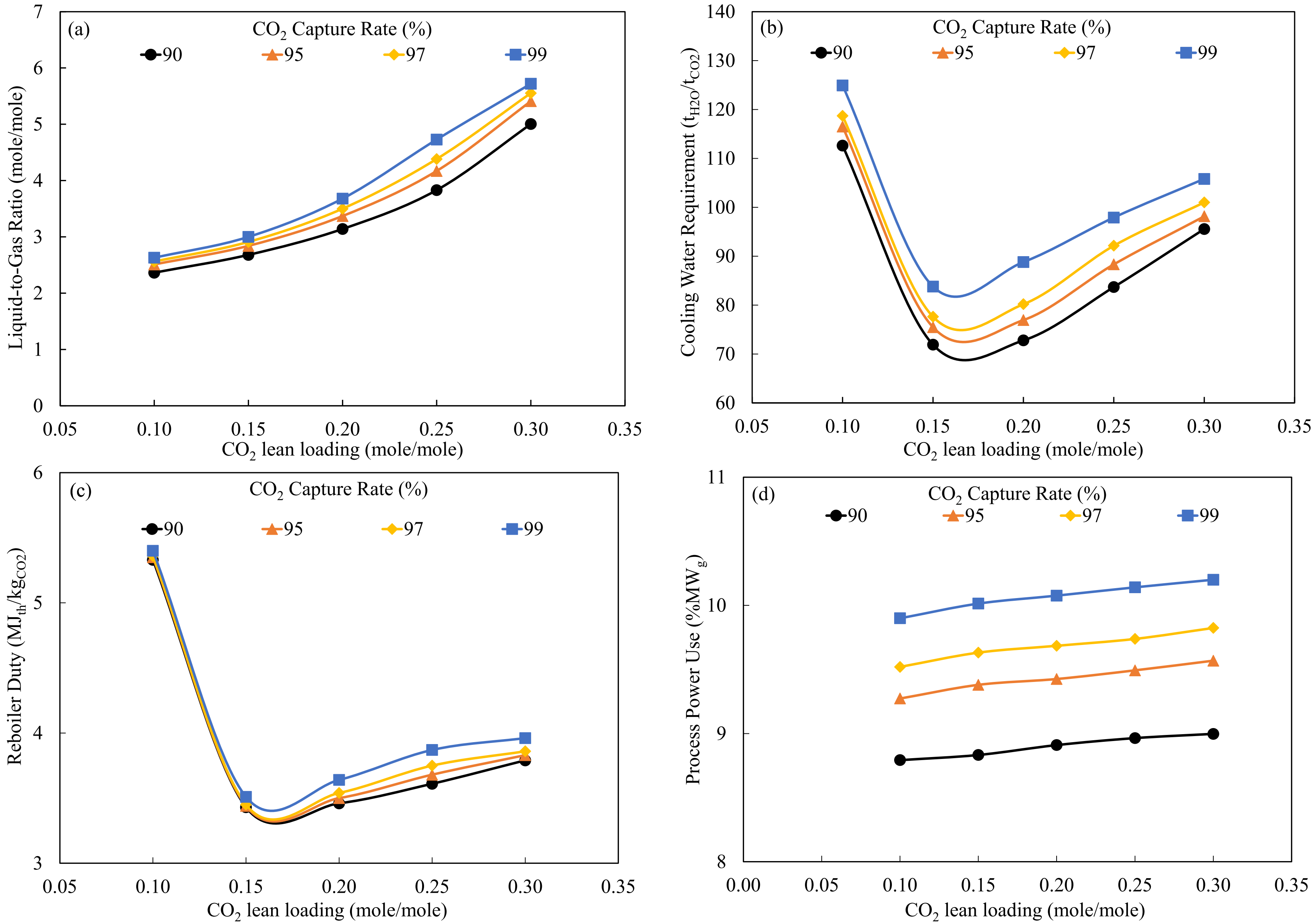
***Figure 4**: Effects of inlet flue gas temperature on key performance parameters of coal power plants: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*

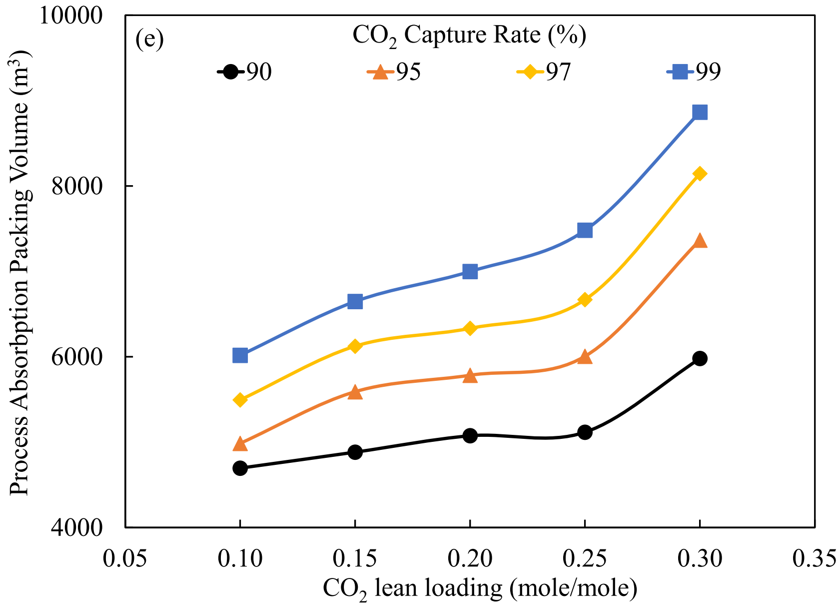
For a given capture rate, the reboiler duty is similar for all the flue gas temperatures, as shown in Figure 4(c). The power requirement increases with the flue gas temperatures and capture rates. This is because at higher temperatures and capture rates, a higher solution flow rate is needed, and hence the blower and solution pumping rise as shown in Figure 4(d). Additionally, a high capture rate means more CO_2 will pass through the compressors, as a result, the overall power requirement rises. When the CO_2 capture rates increase from 90% to 99%, the power use increases by 11.5% on average for all the flue gas temperatures. Likewise, increasing the flue gas temperature from 30 to 50 °C elevates the power use by 11% on average for all the capture rates.

The process absorption and stripping packing volumes for different capture rates are shown in Figure 4(e, f). The effect of inlet flue gas temperature on the absorber diameter is minor. When the CO_2 capture rates increase from 90% to 99%, the process stripping packing volume increases by 21.5% on average for all the flue gas temperatures. Likewise, increasing the inlet flue gas temperatures from 30 to 50 °C elevates the stripping packing volume by 6.3% on average for all the capture rates.

2.6.1.4 Effects of CO_2 Lean Loading

Figures 5(a) and 5(b) show the effects of CO_2 lean loading on the L/G ratio and the cooling water requirements of the capture system. The L/G ratio increases by 12%, 15%, 19%, and 24% as the lean loading rises from 0.10 to 0.15 mole/mole, from 0.15 to 0.20 mole/mole, from 0.20 to 0.25 mole/mole, and from 0.25 to 0.30 mole/mole, respectively. Lean loading has complicated effects on the cooling water requirement. Low lean loading (0.10 mole/mole) stripper condenser requires more cooling water, mainly due to the higher vapor production in the stripper than that of medium lean loading (0.15 and 0.20 mole/mole). The water requirement again increases as the lean loading rises higher lean loading beyond 0.20 mole/mole. It is due to the higher flow rate of treated flue gas to the wash column and the higher lean solution flow rate to the cooler.


(a)
CO2 Capture Rate (%)
90
95
97
99
Liquid-to-Gas Ratio (mole/mole)
CO2 lean loading (mole/mole)
(b)
CO2 Capture Rate (%)
90
95
97
99
Cooling Water Requirement (tH2O/tCO2)
CO2 lean loading (mole/mole)
(c)
CO2 Capture Rate (%)
90
95
97
99
Reboiler Duty (MJth/kgCO2)
CO2 lean loading (mole/mole)
(d)
CO2 Capture Rate (%)
90
95
97
99
Process Power Use (%MWg)
CO2 lean loading (mole/mole)



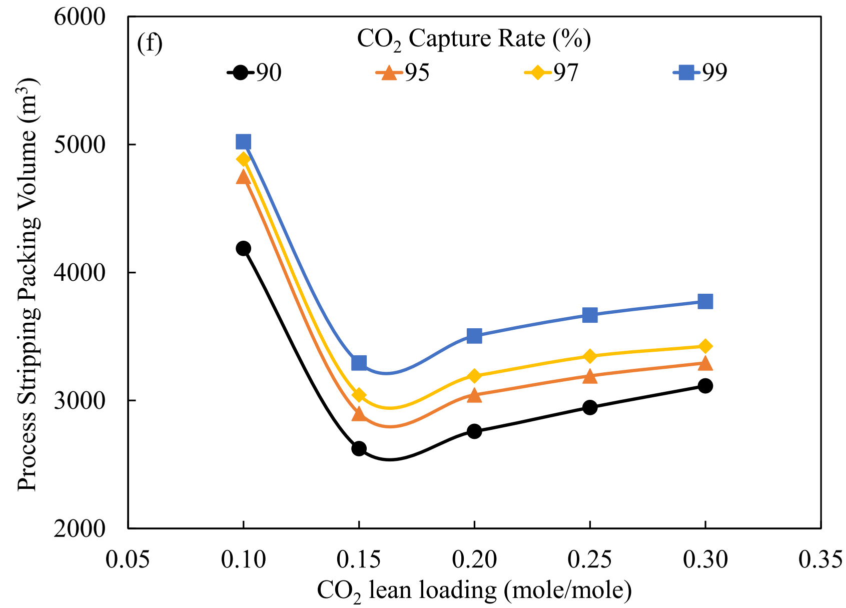


***Figure 5**: Effects of CO_2 lean loading on key performance parameters of coal power plants: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*

Like the cooling water requirement, a low lean loading (0.10 mole/mole) requires higher reboiler duty than that of medium and high lean loading, as more steam is necessary to regenerate the low CO_2 loaded solvent. At high lean CO_2 loadings (0.25 and 0.30 mole/mole), the solvent circulation flow rate is higher than the other lean loadings, leading to an increase in sensible heat, as shown in Figure 5(c). The lean loading values of 0.15 to 0.20 are energetic optimum, as the reboiler duties are lower at these levels. When the CO_2 capture rates increase from 90% to 99%, the process power use increases by 12% on average.

The effects of CO_2 lean loading on the process absorption and stripping packing volumes are shown in Figure 5(e, f). When the CO_2 capture rates increase from 90% to 99%, the process absorption packing volume increases by 28% and the process stripping packing volume increases by 19% on average for all the lean loadings. The process absorption and stripping packing volumes increase with the lean loadings and capture rate to accommodate higher CO_2 concentrations and solution flow rates. The exception is only for the stripping packing volume for 0.10 lean loading. The process of stripping packing volume for a 0.10 lean loading is higher due to the higher vapor production in the stripper, which requires a wider diameter.

2.7. Base Case Assumptions of NGCC Power Plants

Similar to the base case of CO_2 capture from the flue gas of a PC power plant, a baseline scenario involving two turbines from an NGCC power plant is established to assess various CO_2 capture strategies. Table 8 shows a summary of the inlet flue gas characteristics of the 2 turbines NGCC power plant. The properties of lean solution stream, heat exchanger, solvent circulation pumps, wash column, lean solution cooler, stripper specifications, and CO_2 compression unit are the same as the base case of CO_2 capture from a coal-fired power plant. The absorber's number of stages is increased to 60 to improve removal efficiency. Each absorber is equipped with one intercooler.

Table 8: Base Case Assumptions of CO_2 Capture from NGCC-fired Power Plant with 2 Turbines.

Parameter	Value			
Plant Size (MW_g)	591.7			
Inlet Flue Gas				
Flow Rate (tonne/hr)	3103			
Composition (mole%)	CO_2 4.35	H_2O 8.4	O_2 11.6	N_2 75.65
Temperature (°C)	45			
Pressure (bar)	1.10			
CO_2 Capture Rate (%)	90, 95, 96, 97, and 98			

2.8. Results and Discussion of NGCC Power Plants

An NGCC power plant with 2 turbines having 591.7 MW gross capacity of electricity is considered. It releases flue gas at a flow rate of 3,103 tonne/hr (109,500 kmole/hr) containing 4.35 mole% of CO_2 content. To treat it for the CO_2 removal process, 3 absorbers and 1 stripper are used to keep the column approach to flood below 70%. Table 9 shows the input and output summary of major parameters and equipment. The lean solution flow rate and absorber packing volume are varied to achieve the desire capture rates of 90, 95, 96, 97, and 98%. The CO_2 lean loading is fixed at 0.20 mole CO_2/mole MEA. The temperature and pressure of lean solution are assumed to be 40 °C and 1.1 bar, respectively. The L/G ratio, reboiler duty, cooling water requirement, pumps and compressor power requirement, and the design dimensions of absorbers and stripper columns increases with the CO_2 capture rates.

Table 9: Summary of Major Parameters and Equipment for Deep Carbon Capture System in NGCC Plants.

Parameter	**Capture Rate (%)**				
	90	**95**	**96**	**97**	**98**
Lean Solution Flow Rate (kmole/s)	36.43	38.96	39.75	41.12	44.39
Liquid-to-Gas Ratio (L/G) (mole/mole)	1.2	1.28	1.31	1.35	1.46
Absorber Column (m) (each)[a]	D=11.6, H=11.8	D=11.7, H=11.9	D=11.8, H=12	D=11.85, H=12.05	D=11.95, H=12.15
Absorber Pressure Drop (kPa)	9.03	8.92	8.69	8.6	8.32
Stripper Column (m) (each)	D=10.1, H=10.3	D=10.3, H=10.5	D=10.4, H=10.6	D=10.5, H=10.7	D=10.8, H=11
Stripper Pressure (bar)	1.9	1.9	1.9	1.9	1.9
Condenser Temperature (°C)	30	30	30	30	30
Pumps Power (MWe)	0.431	0.461	0.47	0.487	0.527
Reboiler Duty (MW_{th})	187.94	200.14	203.62	209.1	217.71
Number of Intercoolers	3	3	3	3	3
Absorber Intercooler Cooling Water Requirement (tonne/hr)	1,863	2,196	2,361	2,667	3,411
Wash Column Cooling Water Requirement (tonne/hr)	3,124	3,822	3,976	4,075	3,626
Lean solution Cooler Cooling Water Requirement (tonne/hr)	3,480	3,721	3,823	4,011	4,426
Condenser Cooling Water Requirement (tonne/hr)	3,900	4,191	4,289	4,477	5,260
CO_2 Capture Unit Cooling Water Requirement (tonne/hr)	12,367	13,930	14,449	15,230	16,723
Compression Cooling Water Requirement (tonne/hr)	1,865	1,967	1,989	2,009	2030
Total Cooling Water Requirement (tonne/hr)	14,231 (75.41 tonne	15,897 (79.871 tonne	16,438 (81.66 tonne	17,239 (84.742 tonne	18,753 (91.26 tonne

	H_2O/tonne CO_2)	H_2O/tonne CO_2)	H_2O/tonne CO_2)	H_2O/tonne CO_2)	H_2O/tonne CO_2)
MEA Make-up (kg/hr)	13.18	13.09	13.03	13	12.88
Compressor Power (MWe)	17.62	18.61	18.81	19	19.20
Amount of CO_2 Captured (tonne/hr)	188.68	199.15	201.25	203.36	205.46

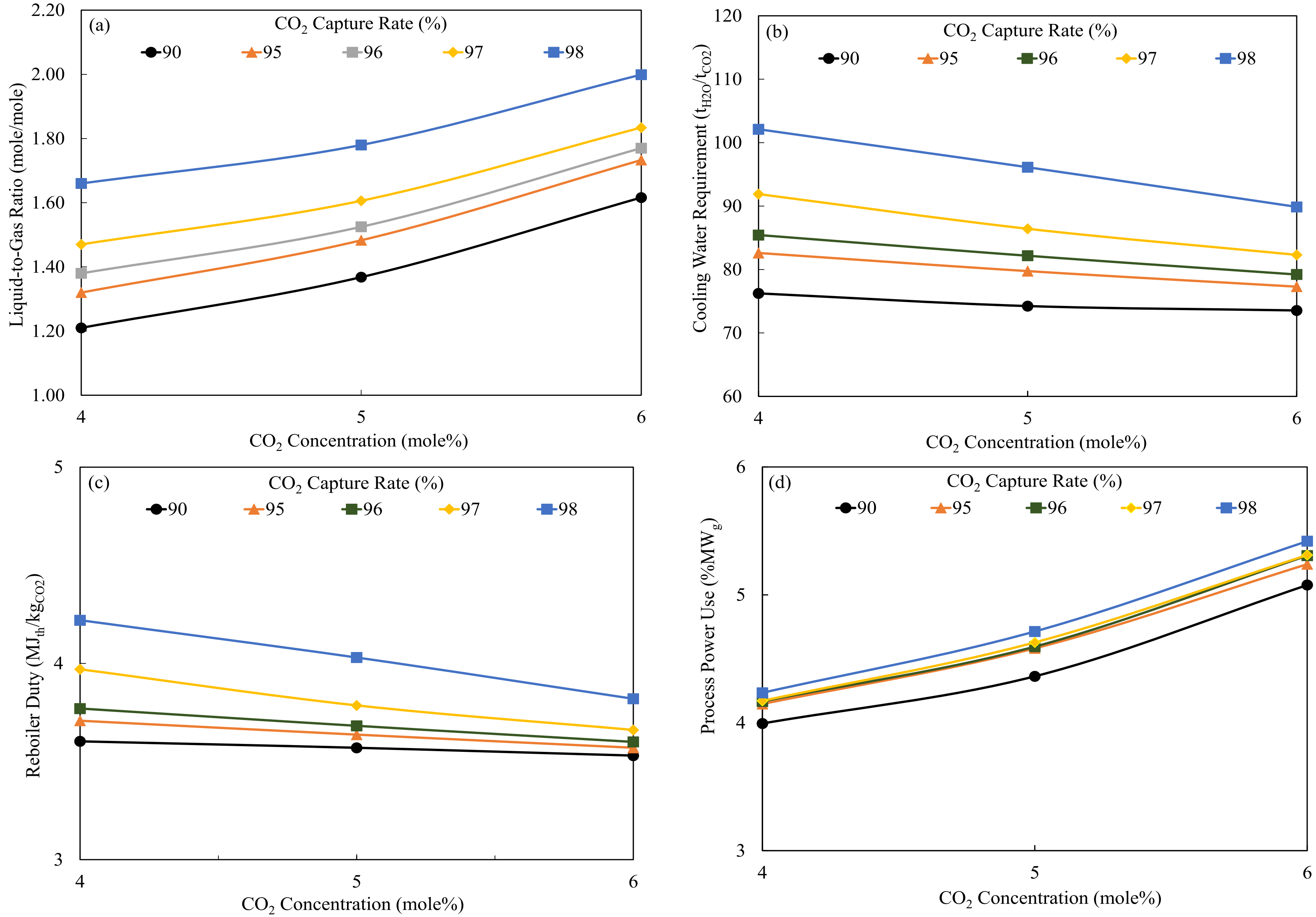
a. D = diameter; H = packing height.

To evaluate the capture process, CO_2 concentration, and flue gas (number of turbines) flow rate are varied to analyze their effect on the performance of the capture process.

2.8.1 Sensitivity Analysis for NGCC Power Plants

This section is dedicated to sensitivity analysis by varying the flue gas flow rate and CO_2 concentration in the flue gas. By configuring the number of turbines, the flue gas flow rates are varied.

The concentration of CO_2 in the flue gas has a significant effect on the CO_2 capture process. CO_2 concentration in the flue gas varies from 4 mole% to 6 mole% to analyze the major performance parameters. Figures 6–10 illustrate the key performance parameters influenced by CO_2 concentrations. They include the L/G ratio, cooling water demand, reboiler duty, process power use, and the required absorption and stripping packing volumes. Figures 6(a)–10(a) demonstrate that the L/G ratio increases with the rise of CO_2 concentration and capture rates. This trend occurs because a higher capture rate necessitates a higher solution flow rate, while the flue gas flow rate remains constant. At a CO_2 concentration of 4 mole% in the flue gas, the L/G ratios for capture rates of 90, 95, 96, 97, and 98% across all turbines are in the following ranges: 1.2–1.21, 1.27–1.32, 1.3–1.38, 1.33–1.47, and 1.41–1.66, respectively. The corresponding averages for these ranges are 1.2, 1.29, 1.32, 1.37, and 1.5. The variation in L/G ratios are due to the differences in the absorption packing volume used in the CO_2 capture processes from the flue gases of turbines 1–5. This trend is nearly the same in 5 mole% and 6 mole% concentrations in flue gas. At a 5 mole% CO_2 concentration in flue gas, the average of the L/G ratios for 90, 95, 96, 97, and 98% CO_2 capture rates are 1.36, 1.45, 1.48, 1.53, and 1.64, respectively. Similarly, at a CO_2 concentration of 6 mole%, the corresponding averages are 1.61, 1.71, 1.74, 1.78, and 1.87.


(a)
CO2 Capture Rate (%)
90
95
96
97
98
Liquid-to-Gas Ratio (mole/mole)
2.20
2.00
1.80
1.60
1.40
1.20
1.00
4
5
6
CO2 Concentration (mole%)
(b)
CO2 Capture Rate (%)
90
95
96
97
98
Cooling Water Requirement (tH2O/tCO2)
120
110
100
90
80
70
60
4
5
6
CO2 Concentration (mole%)
(c)
CO2 Capture Rate (%)
90
95
96
97
98
Reboiler Duty (MJth/kgCO2)
5
4
3
4
5
6
CO2 Concentration (mole%)
(d)
CO2 Capture Rate (%)
90
95
96
97
98
Process Power Use (%MWg)
6
5
4
3
4
5
6
CO2 Concentration (mole%)

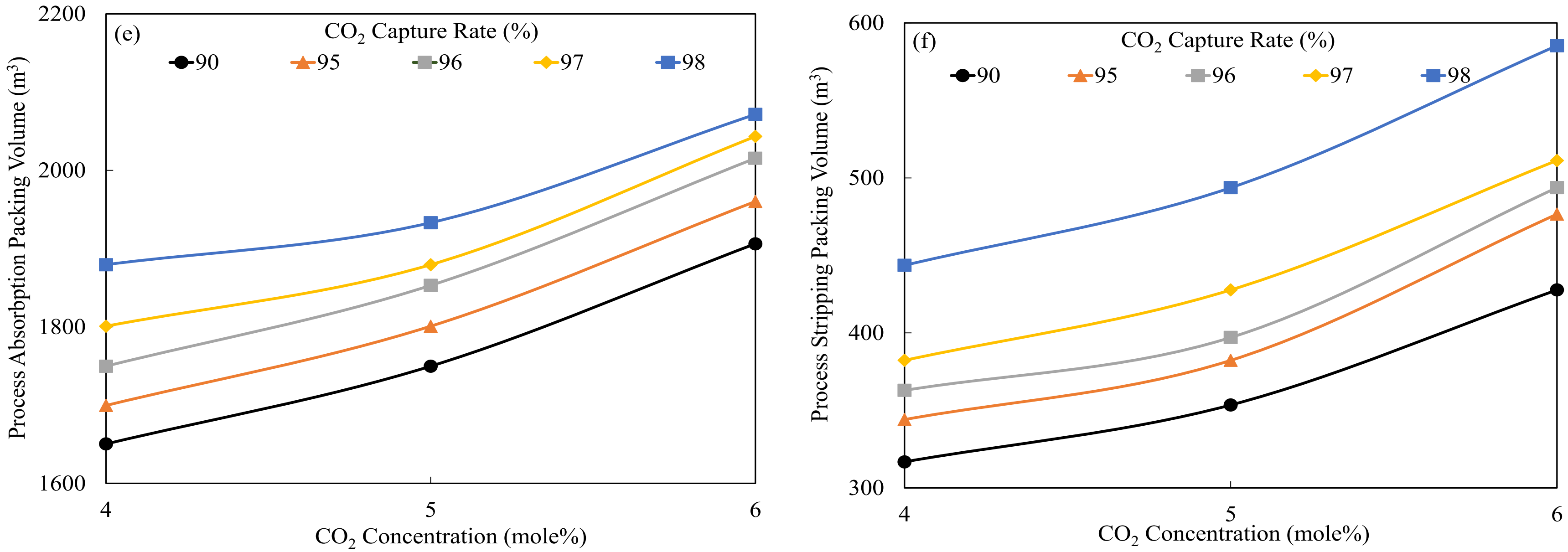
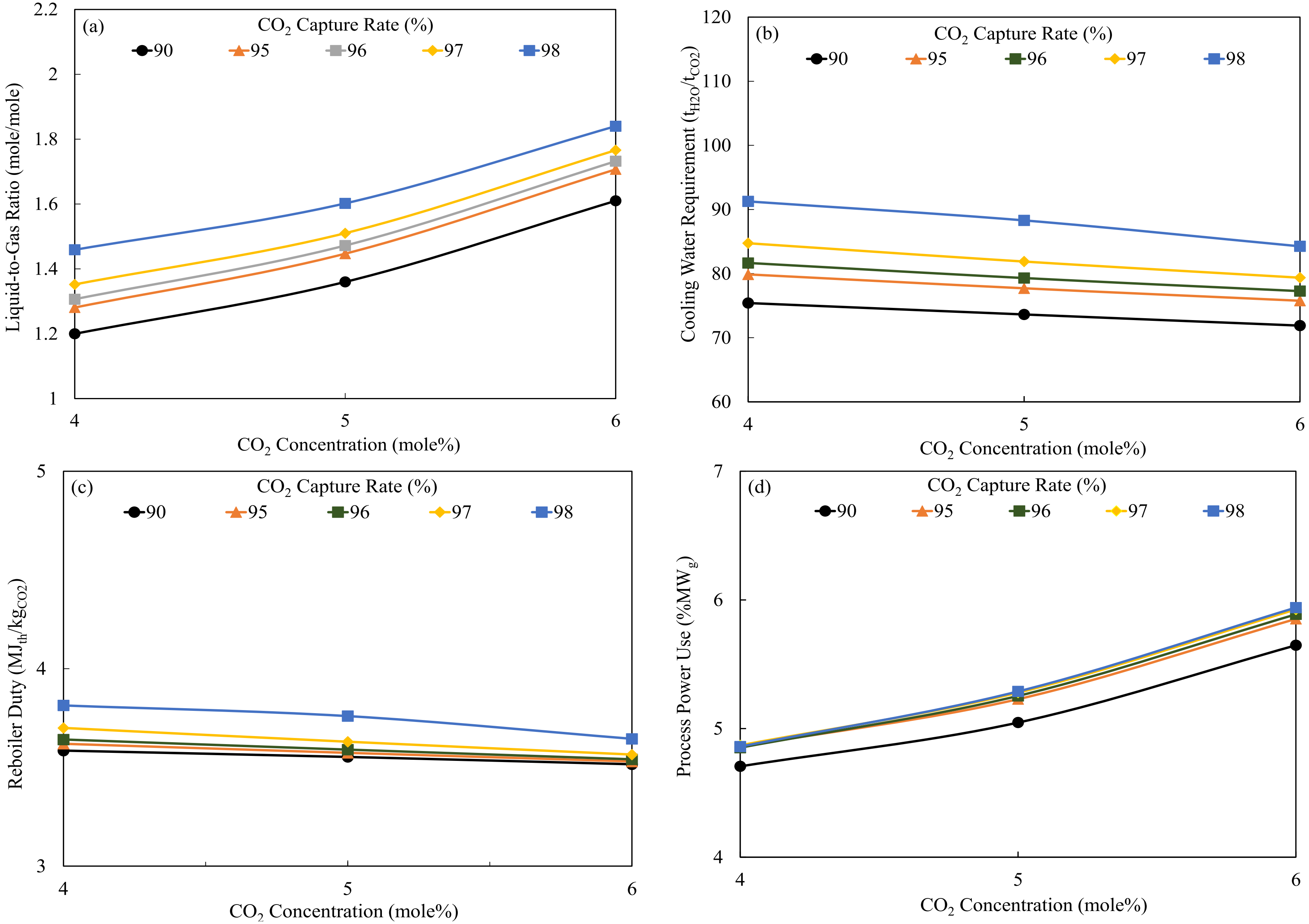


Figure 6: *Effects of CO_2 concentrations on key performance parameters of natural gas power plants with one turbine: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*

Figures 6(b)–10(b) demonstrate that the cooling water requirements increase with higher capture rates but reduce slightly as CO_2 concentrations increase. It is due to the cooling down of the high volume of treated gas in the wash column, the high solution flow rate in the cooler, and the elevated flow rate of CO_2 in the stripper condenser and compressors' intercoolers. At a CO_2 concentration of 4 mole% in flue gas, the average cooling water requirements for 90, 95, 96, 97, and 98% capture rates across all the turbines are 75.6, 80.3, 82.4, 86.1, and 93.8 tonne H_2O/tonne CO_2, respectively. This trend is slightly reduced in the 5 mole% and 6 mole% concentrations in flue gas. At a CO_2 concentration of 5 mole% in flue gas, the corresponding average water consumption values are 73.7, 77.8, 79.9, 82.8, and 89.8 tonne H_2O/tonne CO_2. Similarly, at a CO_2 concentration of 6 mole%, the respective averages are 72.2, 76.1, 77.6, 79.9, and 85.3 tonne H_2O/tonne CO_2.


(a)
CO2 Capture Rate (%)
90
95
96
97
98
Liquid-to-Gas Ratio (mole/mole)
CO2 Concentration (mole%)
(b)
CO2 Capture Rate (%)
Cooling Water Requirement (tH2O/tCO2)
CO2 Concentration (mole%)
(c)
CO2 Capture Rate (%)
Reboiler Duty (MJth/kgCO2)
CO2 Concentration (mole%)
(d)
CO2 Capture Rate (%)
Process Power Use (%MWg)
CO2 Concentration (mole%)

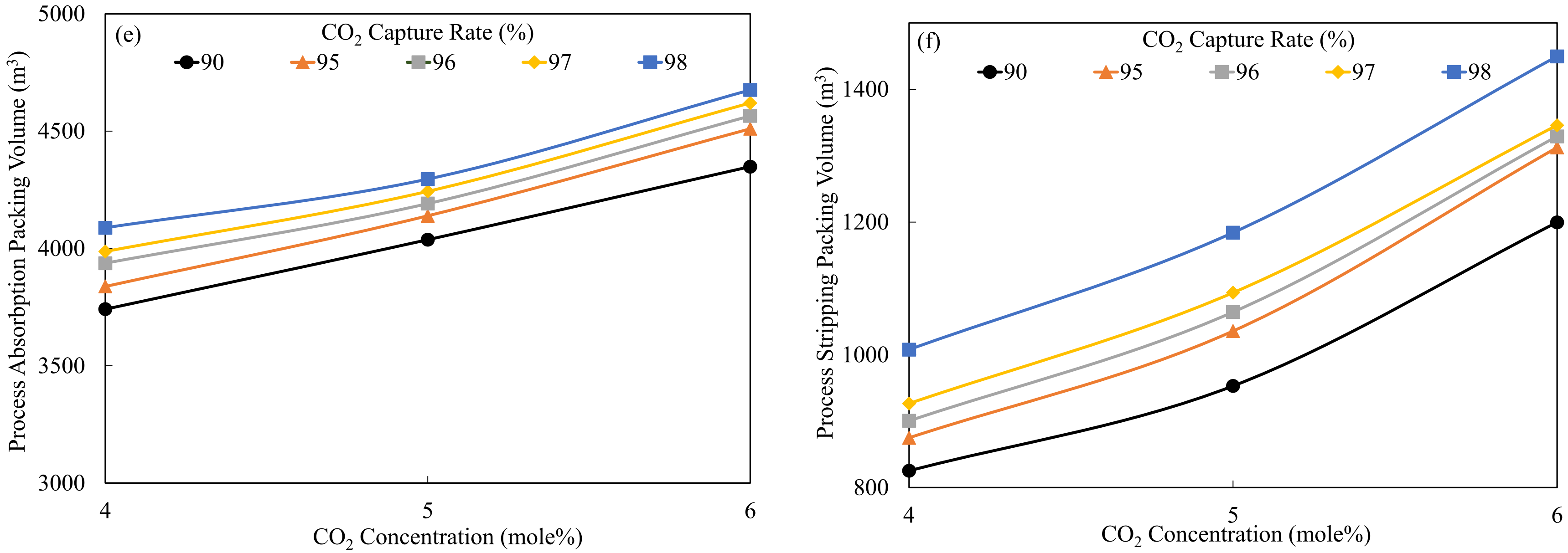
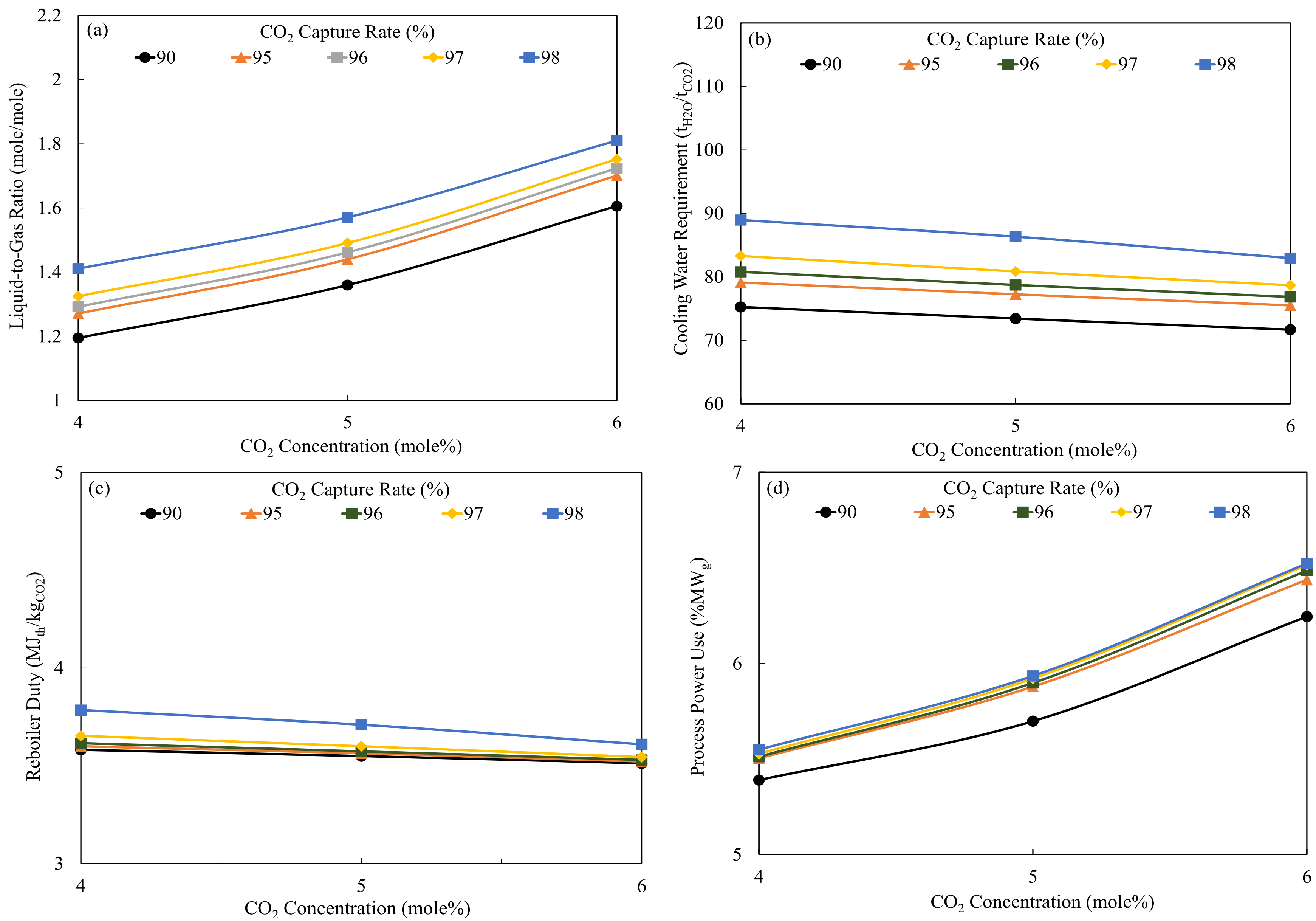


***Figure** 7: Effects of CO_2 concentrations on key performance parameters of natural gas power plants with two turbines: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*

The normalized reboiler duty is almost the same for all CO_2 concentrations, having the same capture rates as depicted in Figures 6(c)–10(c). It only increases with the CO_2 capture rates to remove higher CO_2 gas (95–98%). The reboiler duties of 90, 95, 96, 97, and 98% capture rates from the flue gases of all the turbines are almost 3.59, 3.64, 3.67, 3.75, and 3.91 MJ/kg CO_2, respectively.


(a)
CO2 Capture Rate (%)
90
95
96
97
98
Liquid-to-Gas Ratio (mole/mole)
CO2 Concentration (mole%)
(b)
CO2 Capture Rate (%)
90
95
96
97
98
Cooling Water Requirement (tH2O/tCO2)
CO2 Concentration (mole%)
(c)
CO2 Capture Rate (%)
90
95
96
97
98
Reboiler Duty (MJth/kgCO2)
CO2 Concentration (mole%)
(d)
CO2 Capture Rate (%)
90
95
96
97
98
Process Power Use (%MWg)
CO2 Concentration (mole%)

=

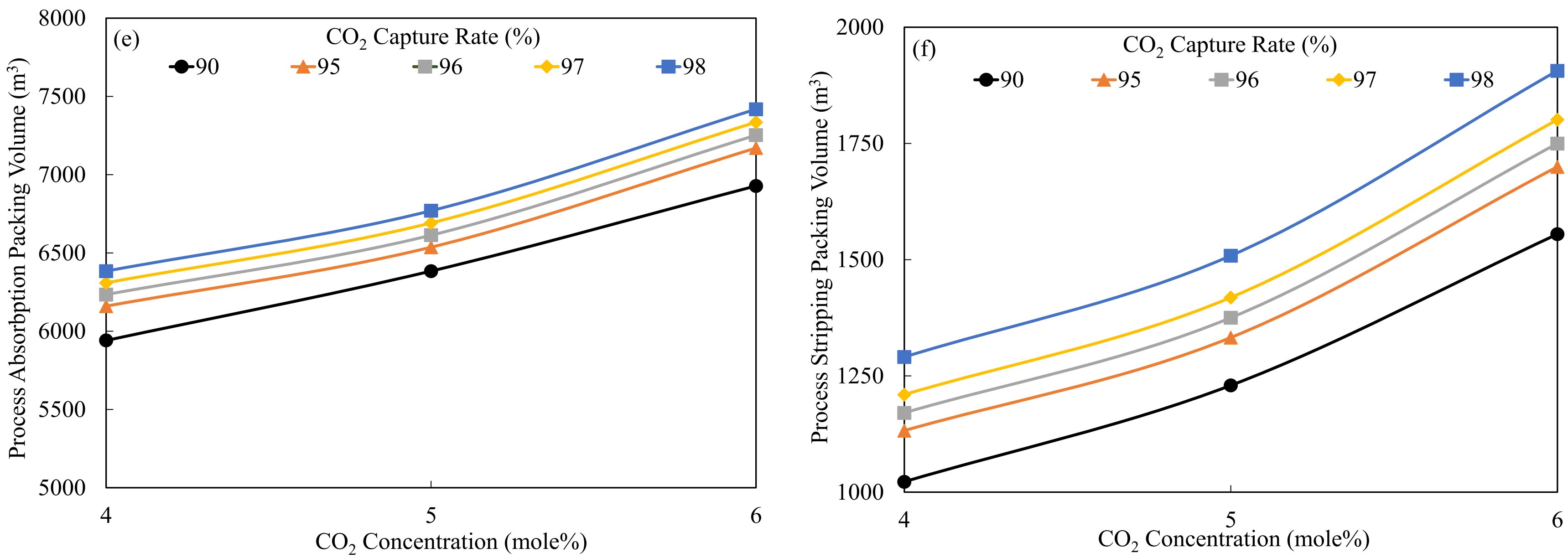
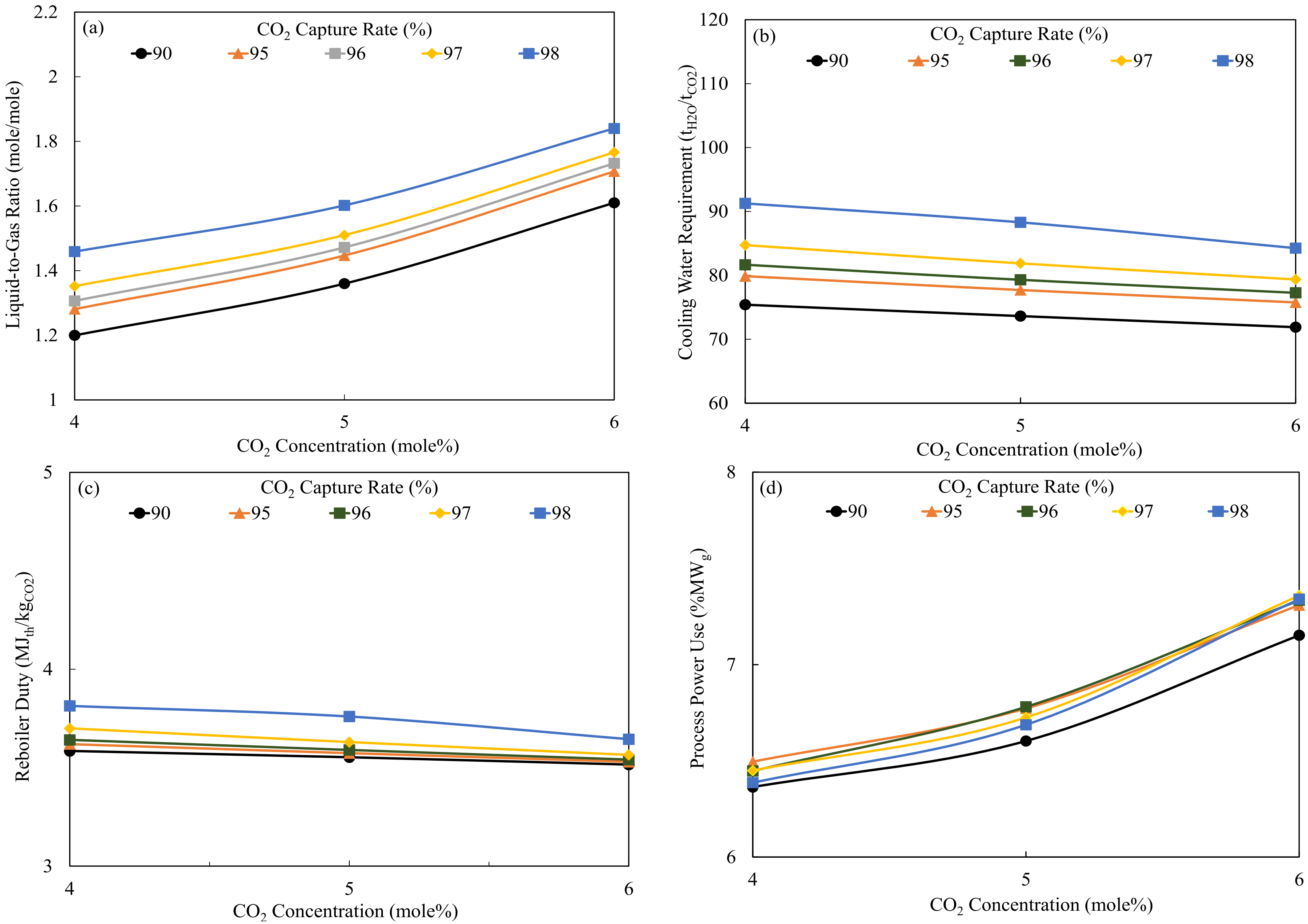


Figure 8*: Effects of CO_2 concentrations on key performance parameters of natural gas power plants with three turbines: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*

The process power requirement is the sum of the power used by the blower, pumps, and compressors. Figures 6(d)–10(d) illustrate that the power requirement increases with CO_2 concentrations and capture rates. In the higher CO_2 capture rates, the solution flow rate is higher, and hence the solution pumping rises. Additionally, a high capture rate means more CO_2 passes through the compressors, thus increasing the overall power requirement. The lowest power use, approximately 4 $\%MW_g$, occurred in the scenario of 90% CO_2 capture from a single turbine with a CO_2 concentration of 4 mole%. In contrast, the highest power use, exceeding 8 $\%MW_g$, is observed in 98% CO_2 capture rate from the flue gas of five turbines with a CO_2 concentration of 6 mole%.


(a)
CO2 Capture Rate (%)
90
95
96
97
98
Liquid-to-Gas Ratio (mole/mole)
CO2 Concentration (mole%)
(b)
Cooling Water Requirement (tH2O/tCO2)
(c)
Reboiler Duty (MJth/kgCO2)
(d)
Process Power Use (%MWg)

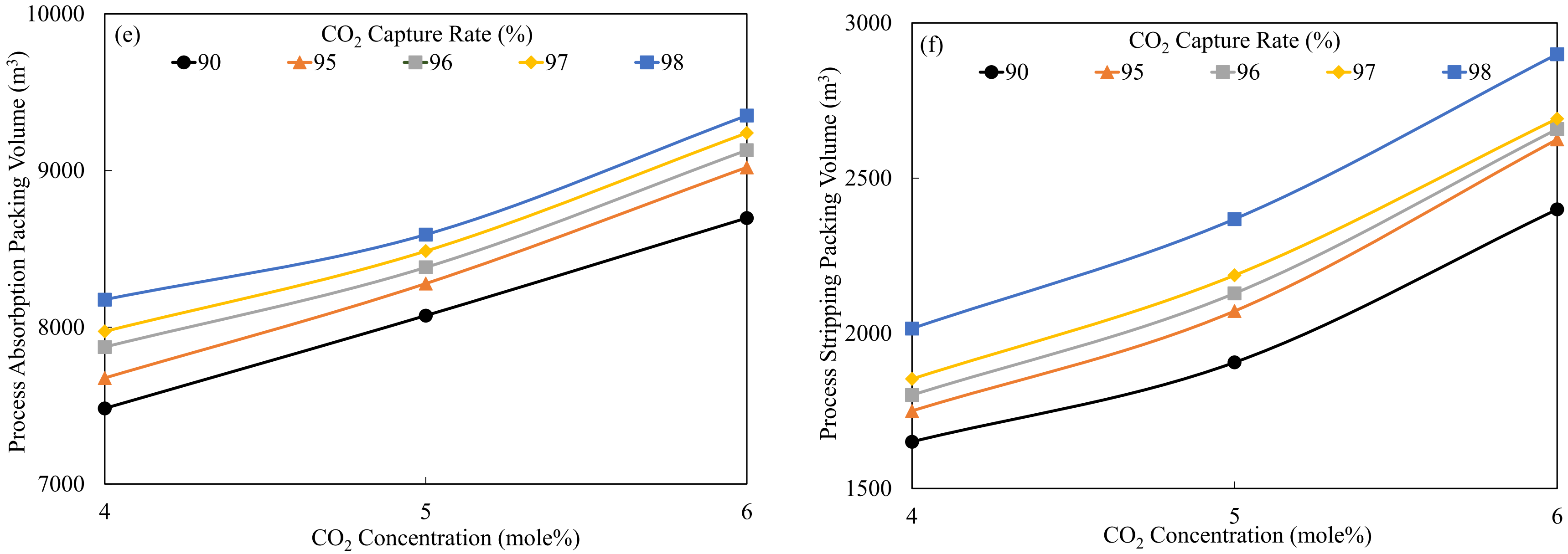
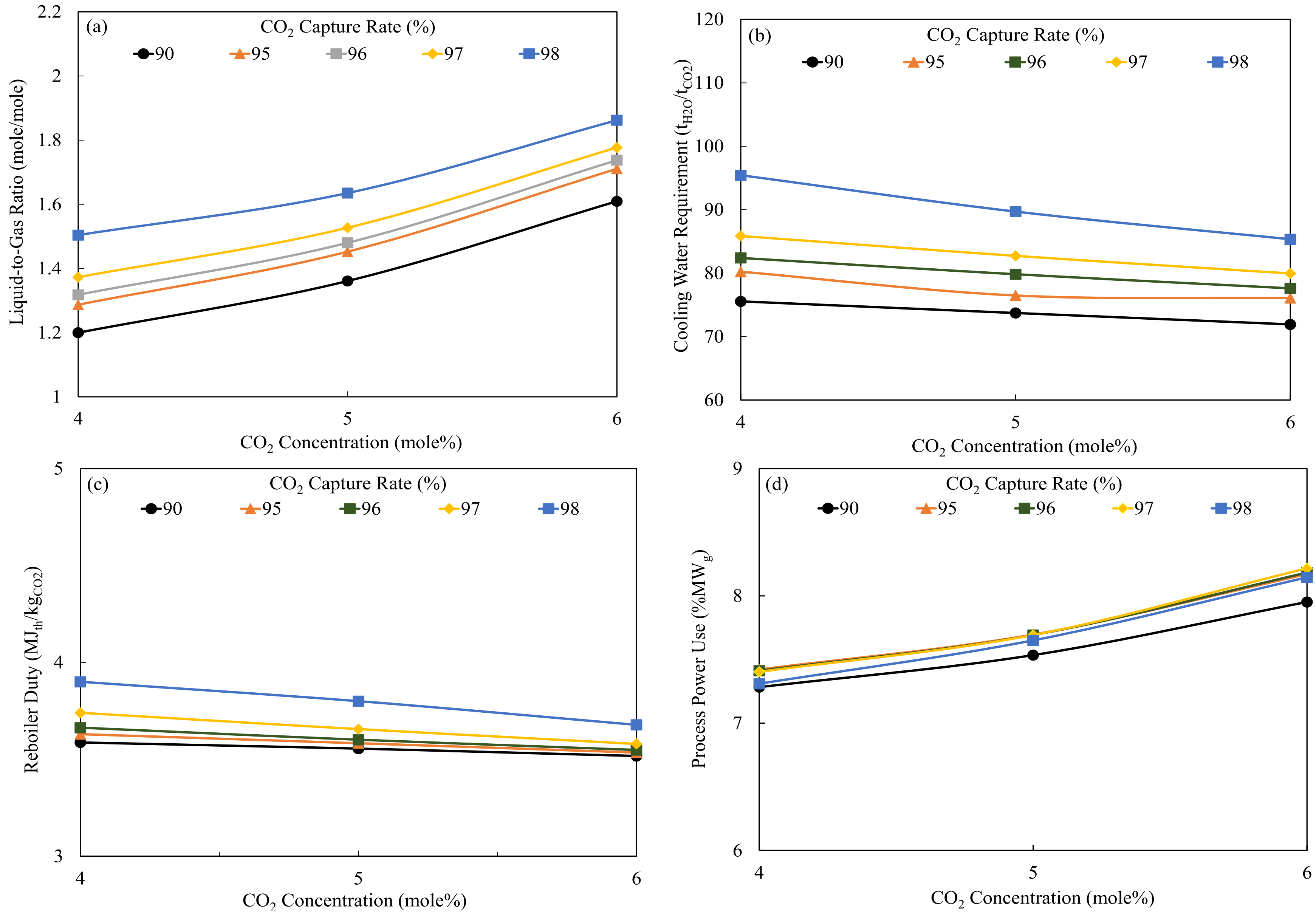


Figure 9: *Effects of CO_2 concentrations on key performance parameters of natural gas power plants with four turbines: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*


(a)
CO_2 Capture Rate (%)
90
95
96
97
98
Liquid-to-Gas Ratio (mole/mole)
CO_2 Concentration (mole%)
(b)
CO_2 Capture Rate (%)
Cooling Water Requirement (t_{H2O}/t_{CO2})
CO_2 Concentration (mole%)
(c)
CO_2 Capture Rate (%)
Reboiler Duty (MJ_{th}/kg_{CO2})
CO_2 Concentration (mole%)
(d)
CO_2 Capture Rate (%)
Process Power Use ($\%MW_g$)
CO_2 Concentration (mole%)

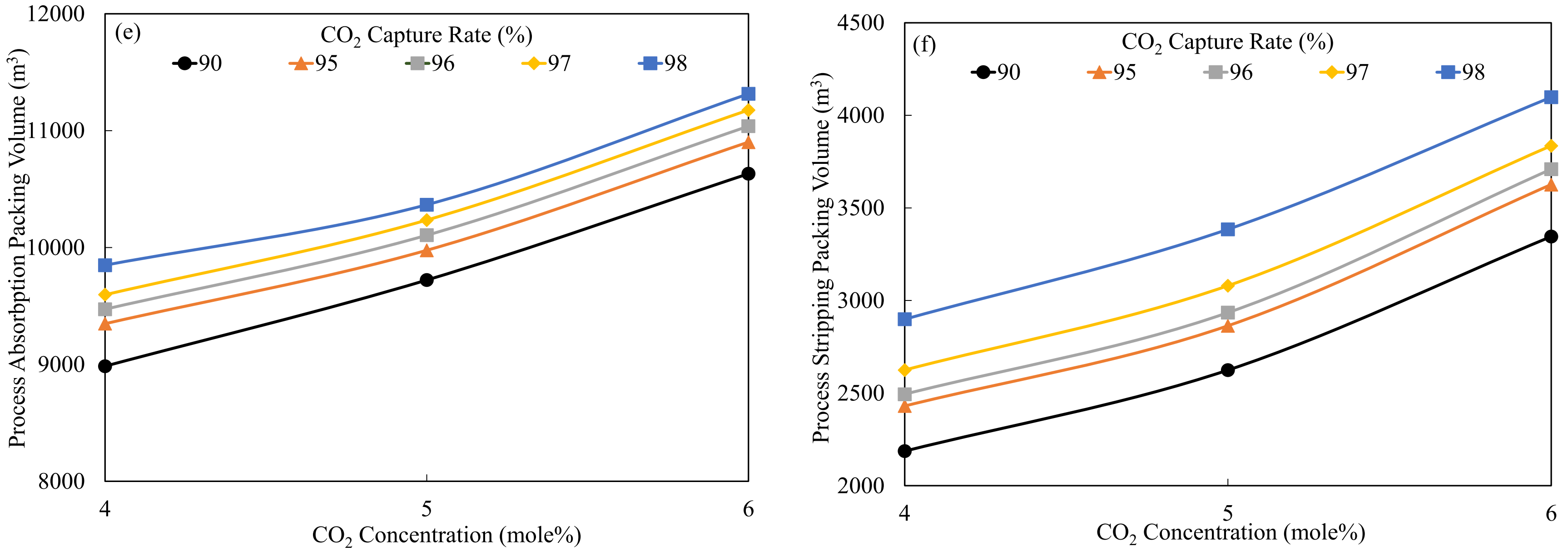


Figure 10*: Effects of CO_2 concentrations on key performance parameters of natural gas power plants with five turbines: (a) L/G ratio, (b) cooling water requirement, (c) reboiler duty, (d) power use, (e) process absorption packing volume, and (f) process stripping packing volume.*

Figure 6(e)–10(e) show that the process absorption volume increases with the CO_2 concentration and capture rates to accommodate higher CO_2 concentrations and solution flow rates. The lowest absorption packing volume, 1,650.4 m^3, is required for the scenario of 90% CO_2 capture from a single turbine with a CO_2 concentration of 4 mole%. In contrast, the highest absorption packing volume, 11,315 m^3, is needed for 98% CO_2 capture from five turbines with a CO_2 concentration of 6 mole%. Like the absorption packing volume, the stripping packing volume (Figures 6(f)–10(f)) increases with the CO_2 concentration and capture rates.

3. Reduced-Order Performance Model Development

There are several parameters that impact gas-liquid equilibrium in CO_2 removal efficiency. This includes the flow rates of flue gas and sorbent, temperature, pressure, flue gas composition, MEA concentration, equipment design, etc. One may classify the absorption of CO_2 in an alkaline media as a first-order process. Thus, a higher CO_2 concentration increases the absorption system's efficiency. MEA has a strong affinity for CO_2, even at low CO_2 concentrations. CO_2 is far more soluble in MEA than it is in many other traditional solvents. This section includes the tool used for simulation, major parameters ranges, explanation of performance parameters, and performance equations.

3.1. Process Simulation Tool

Aspen Plus is used for the simulation of CO_2 capture and compression processes. It is a powerful process engineering tool for the design and steady-state simulation and optimization of process plants.

3.2. Aspen Simulation Runs for CO_2 Capture

CO_2 capture systems have several performance parameters that affect the capturing process. The major performance parameters are explained in the next section. They are varied during performing simulations for CO_2 capture process. Table 10 lists the parameters that are varied and their respective ranges.

Table 10: Major Parameters and Value Ranges of CO_2 Capture from both Coal and NGCC Power Plants.

Parameter	**Units**	**Range**
Coal-fired Power Plants		
CO_2 Concentration in Flue Gas	mole%	7–15
CO_2 Removal Rate	%	90–99
Power Plants Gross Capacity Flue Gas Flow Rate	MW_g tonne/h	400–800 2,033.7–4,068.3
Inlet Flue Gas Temperature	°C	30–50
CO_2 Lean Loading	mole CO_2/mole MEA	0.10–0.30
Flue Gas Flow Rate	tonne/h	2,033.7–4,068.3
Simulation Scenarios (Coal-fired Power Plants)		104
NGCC Power Plants		
CO_2 Concentration in Flue Gas	mole%	4–6
CO_2 Removal Rate	%	90–98
Power Plants Gross Capacity Flue Gas Flow Rate	MW_g tonne/h	296–1479 1,551–7,758
Simulation Scenarios (NGCC Power Plants)		75
Total Number of Simulation Scenarios		179

3.3. CO_2 Capture Performance Parameters

CO_2 capture efficiency (η_{CO_2})

CO_2 capture efficiency, ηco_2, is calculated by dividing the flow rate difference of CO_2 in the flue gas and CO_2 out of the absorber by the flow rate of CO_2 in the flue gas.

MEA concentration (C)

Various amine sorbents have been reported in literature to capture CO_2 from the flue gas. Among them, MEA has the highest rate of absorption and is considered a benchmark solvent. An aqueous MEA solution of 30% MEA and 70% water by weight is used for this purpose. A concentration of MEA higher than 30% increases the degradation rate, while a lower concentration reduces the capture rate.

CO_2 concentration of the flue gas (yco_2)

Flue gases emitted from power plants and industries have different amounts of CO_2 concentration. For example, the flue gas CO_2 concentration of coal-fired power plants is higher than that of natural gas-fired power plants. This parameter is varied during the simulation process in the range of 4–15 mole%.

CO_2 lean loading (ϕ_{Lean})

CO_2 lean loading refers to the mole of CO_2 in the lean solution to the mole of MEA in it. In the stripper column, the MEA solvent is not totally regenerated and contains some “leftover” CO_2. Thus, the lean solvent (regenerated) on recycling back to the absorber column contains CO_2. It is one of the important performance parameters and has significant effects on the regeneration energy requirement, cooling water requirement, and absorber column size.

Liquid-to-gas ratio (L/G)

This ratio is calculated as dividing the liquid (MEA + water solution entering the absorber) molar flow rate by the flue gas molar flow rate in the absorber column. Like CO_2 lean loading, L/G affects the CO_2 capture system significantly. In addition to other parameters, this value needs to be adjusted to get the desire level of capture efficiency.

Flue gas inlet temperature to the absorber ($T_{fg,in}$)

It is the temperature of the flue gas entering the absorber. The desirable value of this parameter for the post-combustion amine-based CO_2 capture process is 40 °C to minimize the absorbent and moisture loss in the exhaust gases [21].

Process absorption (A_p) and stripping packing volumes (S_p)

The process absorption and stripping packing volumes are calculated using Equations 2 and 3:

A_p = number of absorber(s) * (Packing height * πr^2) (2)

S_p = number of stripper(s) * (Packing height * πr^2) (3)

where,

$A_{p=}$ absorber packing volume (m^3)

$S_{p=}$ absorber packing volume (m^3)

r= radius of the columns

Absorber intercooler (Q_{AIC})

The function of the absorber intercooler is to reduce the generated heat by the exothermic reaction between MEA and CO_2 and improve the absorption rate. The intercoolers helped improve the absorption rate and reduce the solution temperature in the column. They collect the solution from the upper section, cool it, and send it to the lower section. Each absorber is equipped with one intercooler, except for the absorber used in the 99% CO_2 capture cases, which had two intercoolers. This design is necessary because the 99% CO_2 capture cases require a higher cooling duty to remove a greater amount of CO_2.

Wash column duty (Q_{AIC})

It is the amount of cooling duty required to avoid the emission of MEA solvent to the environment. Cooling water is required in the wash column.

Total regeneration heat or reboiler duty (Q)

A reboiler is used to regenerate the solvent through stripping heat. Total regeneration or stripping heat is the amount of heat energy required to regenerate the solvent in the reboiler. It is the sum of heat of CO_2 desorption, sensible heat, and heat of water vaporization. High energy requirement is one of the major challenges to the post-combustion chemical solvent absorption process.

Reboiler duty per mole of liquid (Q/L)

This parameter can be obtained by dividing the reboiler duty required for the solvent regeneration into the liquid molar flow rate. This parameter is a function of the lean CO_2 loading, L/G ratio, CO_2 capture efficiency, CO_2 concentration, and the process absorption and stripping packing volumes.

Cooling water requirement ($\dot{m}_{CWR}$)

It is the amount of cooling water required for the absorber intercoolers, wash columns, condenser, lean solution cooler, compression interstage cooling, and compression after stage cooling. The cooling water system is designed to provide water at 21.11 °C and receive at 32.22 °C. Therefore, we have fixed the inlet and outlet temperatures of cooling water at 21.11 °C (70 °F) and 32.22 °C (90 °F), respectively.

Flue gas pressure drop ($\Delta P_{flue\ gas}$)

The pressure of flue gas drops when it flows upward in the absorber column. It is denoted by $\Delta P_{flue\ gas}$. Flue gas blowers help to overcome this pressure drop.

Blower energy (e_{blower})

Flue gas blowers help to overcome the pressure drop of the flue gas in the absorber column. Blower energy refers to the electrical energy required to elevate the pressure of a unit mass of flue gas stream to overcome the pressure drop.

CO_2 compression energy (e_{comp})

It is the amount of electrical energy required to compress a unit mass of the product CO_2 stream to the target pressure of 15.25 MPa. In addition to the pressure ratio, the flow rate of the product CO_2 stream is a major factor in determining this parameter.

Circulation pumping energy (e_{pump})

It is also the electrical energy required to elevate the pressure of a unit mass of solution streams (rich solution stream to P1 and lean solution stream to P2) to the specified pressure. It mainly depends on the flow rate of the solution stream.

Amine scrubber power use (e_{total})

It is the sum of compression, pumps, and blower electrical energy divided by the gross power out of the power plant.

3.4. Performance Equations

The functional relationships between the different essential performance parameters are defined by the performance equations. Equations for multivariate linear regression are generated using the data from the process simulation model through the Microsoft Excel Regression Analysis tool. The total number of carbon capture cases (observations) from both PC and NGCC power plants is 179, comprising 104 cases from PC plants and 75 cases from NGCC plants. The equation used for estimating the liquid-to-gas (L/G) ratio shows that it depends on the CO_2 concentration in the flue gas, the lean CO_2 loading, the CO_2 removal efficiency, and the inlet flue gas temperature. The L/G ratio also relies on the MEA concentration in the lean solution, but we have fixed it at 30 wt.%, so it is not included in the equation. Absorber intercooling also has a minimal impact on the L/G ratio, but we avoid it to reduce the number of independent variables.

Similarly, the regeneration heat requirement (Q) is a function of the lean CO_2 loading, the product CO_2 flow rate, and the lean solution flow rate to the absorber. The energy requirement increases by increasing the capture rate and lean solution flow rate. A low lean loading (0.10 mole/mole) requires higher reboiler duty than that of medium and high lean loading, as more steam is necessary to regenerate the low CO_2 loaded solvent. At high lean CO_2 loadings (0.25 and 0.30 mole/mole), the solvent circulation flow rate rises, leading to an increase in sensible heat. The lean loading value of 0.20 is suitable, as the reboiler duty is low at this level.

The higher the flue gas flow rate, the greater the volume of lean solutions needed to effectively absorb the desired amount of CO_2. Therefore, the absorption packing depends mainly on the lean solution and flue gas flow rate, and CO_2 removal efficiency. On the other hand, stripping packing volume depends on the regeneration heat requirement. As the amount of CO_2 entering the stripper increases, more energy is needed for the regeneration process. The power requirements of the compressor and pumps have a direct relationship with the respective fluids passing through the compressors and pumps. The cooling water requirement is a function of multiple variables, including the product CO_2 flow rate, lean solution and flue gas flow rates, and the lean CO_2 loading. Additionally, it depends on the temperatures or duties of the absorber intercoolers, reflux condenser, wash column, compression inter-stage and after-stage coolers, and the lean solution cooler. However, since we have fixed the temperatures of these components, they are not included in the equation for the cooling water requirement.

$$L/G = -7.8734 + 0 \cdot 2808 * yco_2 + 14.4647 * \phi_{L_{ean}} + 0.0479 * \eta co_2 + 0.0119 * T_{fg,in} \quad (4)$$

[adj. $R^2 = 0.98$]

$$Q = 783.6104 - 3951.0546 * \phi_{L_{ean}} - 39.989 * \dot{mol}_{CO_2} + 6.6227 * L \quad (5)$$

[adj. $R^2 = 0.98$]

$$T_{fg,out\,Abs} = 41.6532 - 0.0363 * G' + 0.0431 * \dot{mol}_{CO_2} + 2.9488 * yco_2 - 8.456 * L/G + 60.1465 * \phi_{L_{ean}} + 0.0344 * \dot{m}_{WC} \quad (6)$$

[adj. $R^2 = 0.90$]

$$T_{fg,out\,WC} = 43.3236 - 0.2983 * \eta co_2 + 0.5739 * T_{fg,out\,Abs} \quad (7)$$

[adj. $R^2 = 0.98$]

$$A_p = -8991.3128 + 92.5406 * \eta co_2 + 99.6675 * G + 27.7269 * L \quad (8)$$

[adj. $R^2 = 0.97$]

$$S_p = -411.5465 + 5.9238 * Q \quad (9)$$

[adj. $R^2 = 0.98$]

$$\dot{m}_{CWR} = 85172.5796 - 432048.2111 * \phi_{L_{ean}} - 82.7913 * \dot{mol}_{CO_2} + 863.848 * L \quad (10)$$

[adj. $R^2 = 0.97$]

$$\dot{m}_{WC} = 2247.1402 + 0.9563 * G' - 1879.6695 * \phi_{L_{ean}} - 17.06 * \eta co_2 - 51.2315 * yco_2 \quad (11)$$

[adj. $R^2 > 0.99$]

$$Q_{WC} = -668.72 + 0.0186 * \dot{m}_{WC} + 6.5976 * T_{fg,out\,Abs} + 3.2884 * \eta co_2 \quad (12)$$

[adj. $R^2 = 0.91$]

$$E_{pump} = (-0.01856 + 0.01222 * L) * 75/\eta_{pump} \quad (13)$$

[$R^2 > 0.99$]

$$E_{comp} = \left(0.7991 + 0.0885 * \dot{m}_{CO_2}\right) * 80/\eta_{comp} \quad (14)$$

[$R^2 > 0.99$]

$$Q_{AIC,PC} = 622.7322 - 8.2354 * Tfg,\ out\,Abs + 2.7088 * yco_2 - 0.0138 * L' - 1.533 * \eta co_2 + 0.4335 * \dot{m}ol_{CO_2} \quad (15)$$

[adj. $R^2 = 0.90$]

[Observations = 104]

$$Q_{AIC,NGCC} = -42.772 - 1.2385 * Tfg,\ out\,Abs + 3.3526 * yco_2 + 0.0359 * L' + 0.968 * \eta co_2 - 0.4586 * \dot{m}ol_{CO_2} \quad (16)$$

[adj. $R^2 = 0.98$]

[Observations = 75]

$$M_{MEA}^{Fluegas\ Exhast} = 0.1788 + 0.0045 * G' - 0.0003 * L' \quad (17)$$

[adj. $R^2 > 0.99$]

$$\Delta P_{flue\ gas,PC} = 1.50213 + 0.00087 * A_p - 0.00026 * L' + 0.00118 * G' + 0.00993 * Q_{AIC,PC} \quad (18)$$

[adj. $R^2 = 0.91$]

[Observations = 104]

$$\Delta P_{flue\ gas,NGCC} = 0.03554 - 0.00203 * A_p + 0.00042 * L' + 0.00498 * G' - 0.00844 * Q_{AIC,NGCC} \quad (19)$$

[adj. $R^2 > 0.99$]

[Observations = 75]

where,

L = Total lean solution flow rate (kmole/s) (L' = tonne/hr)

G = Total inlet flue gas flow rate to the absorbers (kmole/s) ($G' = \text{tonne/hr}$)

L/G = Ratio of molar flow rates of total lean solution to the total inlet flue gas

Q = Total solvent regeneration heat requirement or reboiler duty (MW_{th} or MJ/s)

ϕ_{Lean} = CO_2 lean loading (mole CO_2/mole MEA)

y_{CO_2} = CO_2 concentration in the inlet flue gas (mole%)

η_{CO_2} = CO_2 capture efficiency (%)

$\dot{m}_{CO_2}$ = Product CO_2 flow rate to the compressor (tonne/hr) ($\dot{mol}_{CO_2}$ = kmole/s)

Q_{AIC} = Absorber intercooler duty (MW_{th})

$T_{fg,in\ Abs}$ = Temperature of the inlet flue gas to the absorber (°C)

$T_{fg,out\ Abs}$ = Temperature of the flue gas leaving the absorber or treated flue gas (°C)

$T_{fg,out\ WC}$ = Temperature of the flue gas leaving the wash column (°C)

A_p = Process absorption packing volume (m^3)

S_p = Process stripping packing volume (m^3)

$\dot{m}_{CWR}$ = Cooling water requirement (tonne/hr)

$\dot{m}_{WC}$ = Flow rate to wash column (tonne/hr)

Q_{WC} = Wash column duty (MW_{th})

E_{comp} = Power requirement for CO_2 compression (MW_e)

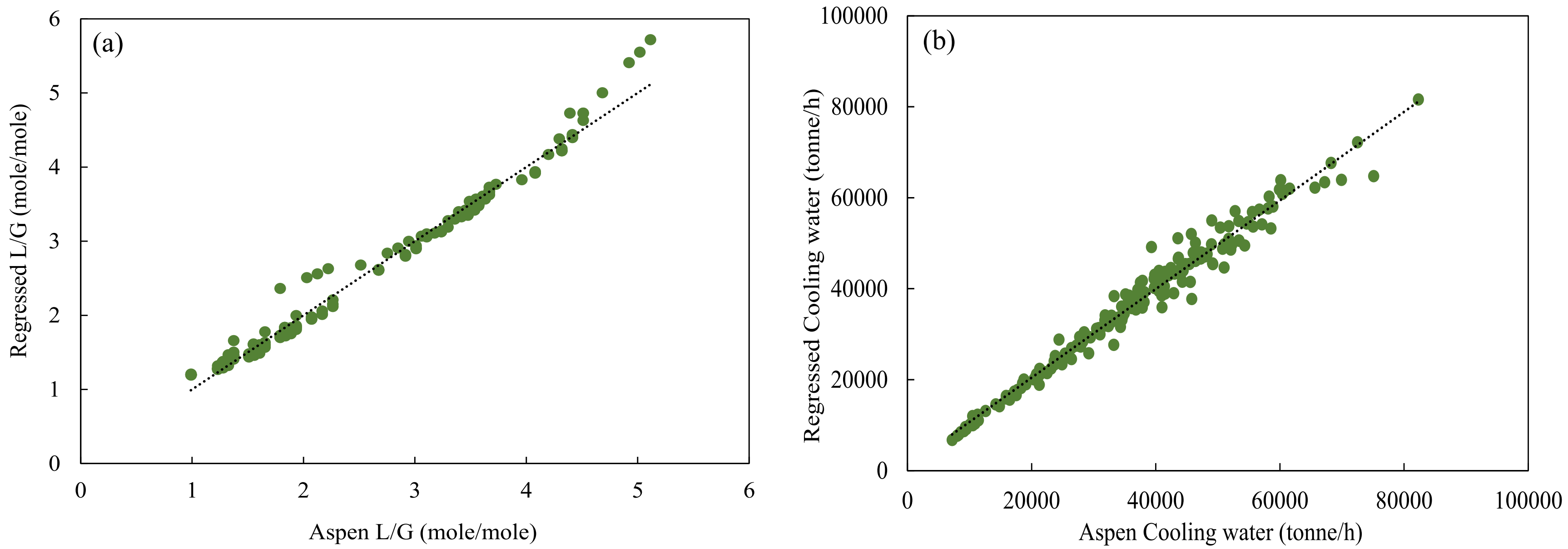
E_{pump} = Solution circulation pumps power (MW_e)

MEA_{makeup} = Sorbent makeup (flue gas exhaust) (kg/hr)

$\Delta P_{flue\ gas}$ = Flue gas pressure drop (kPa)

Absorber intercoolers are employed to lower the column temperature, thereby enhancing the absorption rate. Consequently, they are linked to the absorber top temperature, capture rates, lean solution flow rates, and CO_2 concentration in the flue gas. The primary function of the wash column is to recover solvent emissions. Its performance is influenced by the stream entering it, which originates from the absorber's top, also referred to as vent gases. As the flow rate of the vent stream increases, the duty of the wash column also rises.

Figures 11 shows the relationship between the regressed data (obtained through the above equations) and the Aspen model results of the carbon capture process performance parameters. Appendix A also presents supporting information regarding the regression analysis summaries for all equations, including the L/G ratio, regeneration heat requirement, treated flue gas temperature, wash column outlet temperature, compressor and pump use, intercooler and wash column duties, MEA make-up, absorber pressure drop, cooling water requirement, and absorption and stripping process packing volumes.



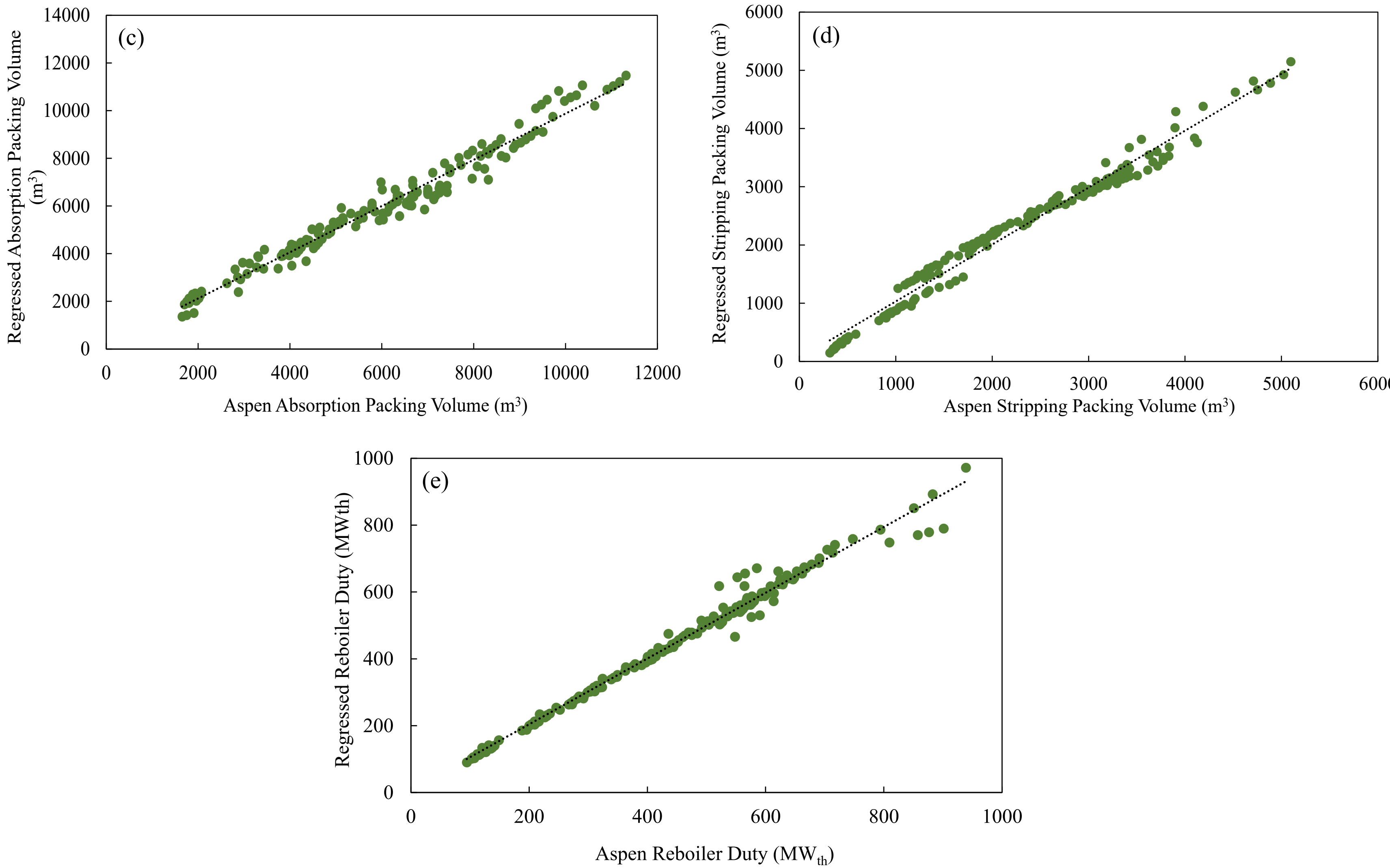


***Figure 11**: Comparison of regressed and Aspen-modeled CO_2 capture data: (a) liquid-to-gas ratios, (b) cooling water requirement, (c) absorption packing volume, (d) stripping packing volume, and (e) regeneration heat requirement.*

4. Cost Model Development

The framework of the economic model developed for the Deep Carbon Capture System follows the framework used in the IECM to ensure consistency in economic calculations. The economic model developed in this study encompasses the capital costs and operating and maintenance (O&M) costs that are directly linked to the reduced-order performance model in Section 3. The metric, levelized cost of CO_2 separation (\$/tonne CO_2 separated), is estimated to evaluate the cost-effectiveness of a carbon capture system design, as shown in Equation 20:

$$\text{Cost of } CO_2 \text{ Separation} = \frac{TCR_{cc} * FCF + FOM_{cc} + VOM_{cc}}{\dot{m}_{CO_2} * CF * HPY} \tag{20}$$

where,

TCR_{cc} = Total capital requirement of CO_2 capture and compression system (\$)

FCF = Fixed charge factor (fraction/yr)

FOM = Fixed O&M cost of CO_2 capture and compression system (\$/yr)

VOM = Variable O&M cost of CO_2 capture and compression system (\$/yr), excluding the cost of CO_2 transport and storage

$\dot{m}_{CO_2}$ = Mass flow rate of CO_2 product (tonne/hr)

CF = Capacity factor (%)

HPY = Total annual operation hours (8,766 hrs/yr)

4.1. Capital Cost

The total capital requirement (TCR) represents the total capital costs required for the Deep Carbon Capture system. It includes the direct capital cost or process facilities cost (PFC) for equipment purchase and installation and indirect capital costs, such as general facilities cost, engineering & home office fees, contingency costs, the interest charge during construction, and owner's cost, as shown in Figure 12. The PFC is estimated as a function of the technical parameters that decide the sizes and capacities of the major equipment, while indirect capital costs are generally estimated as a function of the total PFC or variable O&M costs.

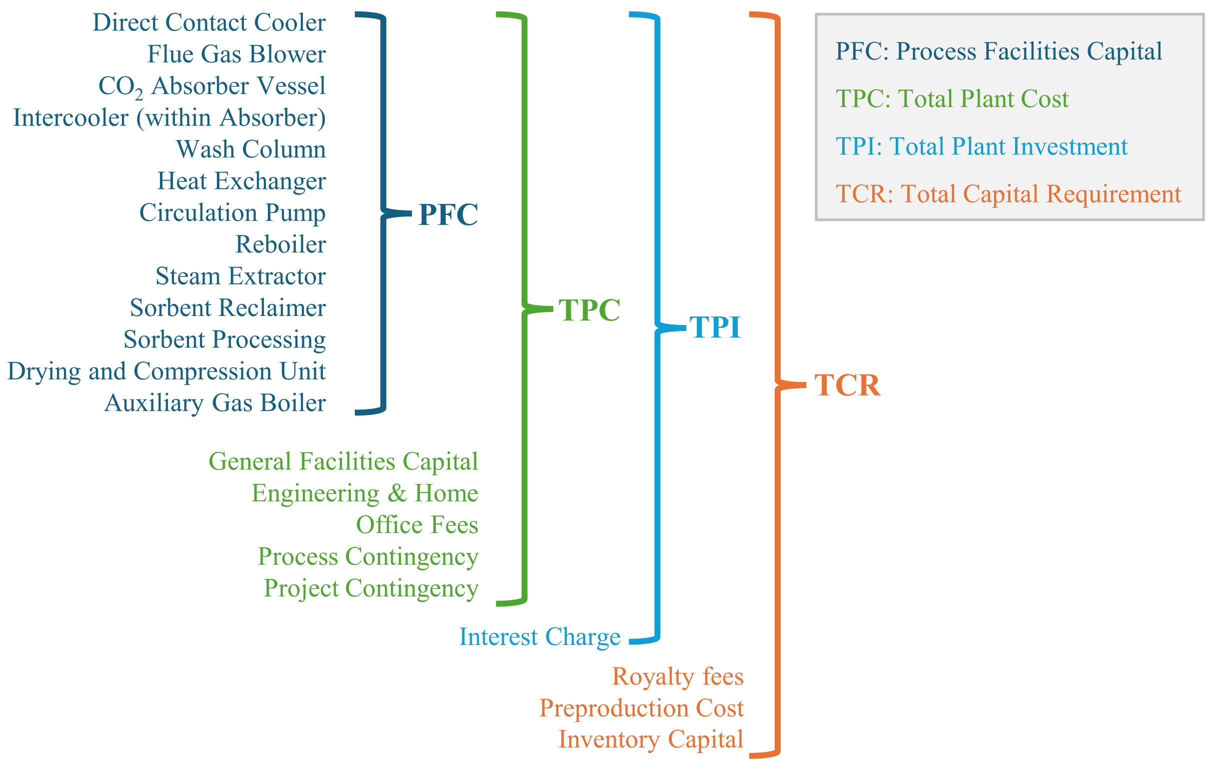


***Figure 12**: Structure of total capital requirement of carbon capture system.*

4.1.1. Process Facilities Costs

Process facilities costs are the purchase and installation costs of process facilities, which include the SO_2 polisher with direct contact cooler, flue gas blower, CO_2 absorber vessel, intercoolers, wash column, heat exchangers, circulation pumps, sorbent regenerator, reboiler, steam extractor, sorbent reclaimer, sorbent processing, and the CO_2 drying and compression unit. The PFC of each component is estimated at the system level and may have multiple trains installed unless otherwise noted.

The PFCs of most components are estimated based on the costs and associated technical information obtained from IECM v11.5, unless otherwise noted. We use IECM v11.5 to model a series of PC and NGCC power plants equipped with an amine-based carbon capture system for 90% CO_2 removal. The PFCs and technical parameters are then fitted to a power function as given by Equation 21. The data points and regression results for PC and NGCC power plants are shown in Appendix B.

$$C_i = a \cdot \left(\frac{X_i}{X_{i,ref}}\right)^b \quad (21)$$

where,

C_i = PFC of process facility, i (M$)

X_i = Scaling parameter

$X_{i,ref}$ = Reference scaling parameter a = regression cost coefficient

b = regression scaling exponent

The reference scaling parameters are obtained from IECM's library, "NETL case 12 (rev2a)", which modeled a PC plant equipped with FG+ Amine-based CO_2 capture system. To be consistent with the NETL's reported costs, the costs in the regressions results are presented in 2007 constant dollars. Additionally, the estimation methods of capital cost of intercoolers, wash columns, and condensers refer to the technical report of the National Renewable Energy Laboratory (NREL) and Pacific Northwest National Laboratory (PNNL) [31]. The annual Chemical Engineering Plant Cost Index (CEPCI) is used to convert the cost into other years when needed.

Table 11: Reference Scaling Parameters of Amine-based CO_2 Capture System.[a]

Parameters	Unit	Value
Flue Gas Flow Rate	tonne/hr	2,957
	kg-mole/hr	102,600
Temperature of Flue Gas Exiting the Direct Contact Cooler	°C	45
CO_2 Concentration	mole%	13.5
Liquid-to-Gas Ratio	mole/mole	3.75
Sorbent Circulation Rate	tonne/hr	9,029
CO_2 Product Flow Rate	tonne/hr	547.4
Regenerator Heat Requirement	kJ/kg CO_2	3,763
Sorbent Makeup Flow Rate	kg/hr	55.25

a. Data source: "NETL case 12 (rev2a)" in IECM library.

The regression results for the process facility costs (PFCs) as functions of performance parameters are presented in Equations 22–35, with nomenclature provided following the equations. Please note that the Equations 25–27, corresponding to intercoolers, wash column, and condensers, are obtained from the technical reports by NREL and NETL [31]. In addition, the capital cost of steam extractors is estimated based on the number of steam extractors, which is determined by the number of absorbers. Each absorber has a maximum train CO_2 capacity of 208.7 tonne/hr (IECM v11.5).

Direct Contact Cooler

$$\mathbf{C_{dc,PC} = 29.64 * \left[\frac{G_{mass}}{G_{mass,ref}} * \frac{T_{fg,in\ Abs}}{T_{fg,in\ Abs,ref}}\right]^{0.8928} \quad [R^2 = 0.99]} \qquad (22a)$$

$$\mathbf{C_{dc,NG} = 24.312 * \left[\frac{G_{mass}}{G_{mass,ref}} * \frac{T_{fg,in\ Abs}}{T_{fg,in\ Abs,ref}}\right]^{0.8752} \quad [R^2 = 0.98]} \qquad (22b)$$

Flue Gas Blower

$$\mathbf{C_{blower,\ PC} = 6.1038 * \left[\frac{G_{mass}}{G_{mass,ref}}\right]^{0.8928} \quad [R^2 = 0.99]} \qquad (23a)$$

(23b)

$$C_{blower,\,NG} = 4.4603 * \left[\frac{G_{mass}}{G_{mass,ref}}\right]^{0.8751} \quad [R^2 = 0.98]$$

CO_2 Absorber Vessel

$$C_{abv,\ PC} = 117.11 * \left[\frac{G_{mole} * yco_2 * (L/G)}{G_{mole,ref} * yco_{2,ref} * (L/G)_{ref}}\right]^{0.981} \quad [R^2 = 0.97] \qquad (24a)$$

$$C_{abv,\ NG} = 461.37 * \left[\frac{G_{mole} * yco_2 * (L/G)}{G_{mole,ref} * yco_{2,ref} * (L/G)_{ref}}\right]^{0.8738} \quad [R^2 = 0.98] \qquad (24b)$$

Absorber Intercoolers

$$C_{abi,\,combined} = N_{IC} * 2.353 * (\frac{Q_{AIC}/N_{AIC}}{Q_{AIC,ref}})^{0.65} * 3.18 \qquad (25)$$

Wash Column Condenser

$$C_{wcc,\,combined} = 0.263 * (\frac{Q_{WC}}{Q_{WC,ref}})^{0.44} * 1.88 \qquad (26)$$

Wash Column KO-Drum

$$C_{wckd,\,combined} = 0.1087 * (\frac{\dot{m}_{wc}}{\dot{m}_{wc,ref}})^{0.65} * 2.0 \qquad (27)$$

Heat Exchanger

$$C_{hx,\,combined} = 7.4856 * \left[\frac{M_{solvent}}{M_{solvent,ref}}\right]^{0.897} \quad [R^2 = 0.99] \qquad (28)$$

Circulation Pump

$$C_{cp,\,Combined} = 15.42 * \left[\frac{L_{mass}}{L_{mass,ref}}\right]^{0.8969} \quad [R^2 = 0.99] \qquad (29)$$

Sorbent Regenerator

$$C_{sr,\,combined} = 56.693 * \left[\frac{\dot{m}_{CO_2} * (L_{mass}/\dot{m}_{CO_2})}{\dot{m}_{CO_2,ref} * (L_{mass,ref}/\dot{m}_{CO_2,ref})}\right]^{0.915} \quad [R^2 = 0.99] \qquad (30)$$

Reboiler

$$\mathbf{C_{rb,\,combined} = 24.309 * \left[\frac{\dot{m}_{CO_2} * HR_{regenerator}}{\dot{m}_{CO_2,ref} * HR_{regenerator,ref}}\right]^{0.882} \quad [R^2 = 0.97]} \tag{31}$$

Steam Extractor

$$\mathbf{C_{se,\,combined} = 1.1065 * \left[Int\left(\frac{\dot{m}_{CO_2}}{M_{CO_2,max}}\right) + 1\right]} \tag{32}$$

Sorbent Reclaimer

$$\mathbf{C_{sr,\,combined} = 1.165 * \left[\frac{MEA_{makeup}}{MEA_{makeup,ref}}\right]^{0.874} \quad [R^2 = 0.99]} \tag{33}$$

Sorbent Processing

$$\mathbf{C_{sp,\,combined} = 1.3161 * \left[\frac{MEA_{makeup}}{MEA_{makeup,ref}}\right]^{0.6} \quad [R^2 = 1.0]} \tag{34}$$

CO_2 Drying and Compression Unit

$$\mathbf{C_{CO2\_compr,\,combined} = 0.0519 * \dot{m}_{CO_2} + 25.943 \quad [R^2 = 0.999]} \tag{35}$$

where,

$C_{i,\,combined}$ = Process facility cost of component i, of combined case (meaning can be applied to both PC and NGCC plants) in M$

$C_{i,\,pc}$ = Process facility cost of component i, that only applies to PC plant in M$

$C_{i,\,NG}$ = Process facility cost of component i, that only applies to the NGCC plant in M$

G_{mass} = Total inlet flue gas mass flow rate to the absorbers (tonne/hr)

$G_{mass,ref}$ = Reference inlet flue gas mass flow rate to the absorbers (2,957 tonne/hr)

$T_{fg,in\ Abs}$ = Temperature of the inlet flue gas to the absorber (°C)

$T_{fg,in\ Abs,ref}$ = Reference temperature of the inlet flue gas to the absorber (°C)

G_{mole} = Molar flow rate of inlet flue gas (kg-mole/hr) to the system

$G_{mole,ref}$ = Reference molar flow rate of inlet flue gas (102,600 kg-mole/hr)

yco_2 = CO_2 concentration of inlet flue gas (mole%)

$yco_{2,ref}$ = Reference CO_2 concentration of inlet flue gas (13.5 mole%)

L/G = Liquid-to-gas ratio of the amine scrubber (mole/mole) at a system level

L/G_{ref} = Reference liquid-to-gas ratio (3.75)

N_{AIC} = Number of intercoolers

Q_{AIC} = Total cooling duty of intercoolers (MW_{th})

$Q_{AIC,ref}$ = Reference cooling duty of an intercooler (37.4 MW_{th})

Q_{WC} = Cooling duty of the condenser of the wash column (MW_{th})

$Q_{WC,ref}$ = Reference cooling duty of a condenser of a wash column (22.73MW_{th})

$\dot{m}_{WC}$ = Mass flow rate of gas entering the wash column (tonne/hr)

$\dot{m}_{WC,ref}$ = Reference mass flow rate of gas entering the wash column (195.8 tonne/hr)

L_{mass} = Mass flow rate of lean solution (tonne/hr)

$L_{mass,ref}$ = Reference mass flow rate of lean solution (9,029 tonne/hr)

$\dot{m}_{CO_2}$ = Mass flow rate of CO_2 product at a system level (tonne/hr)

$\dot{m}_{CO_{2,ref}}$ = Reference mass flow rate of CO_2 product (547.4 tonne/hr)

$HR_{regenerator}$ = Regenerator heat requirement (kJ/kgCO_2)

$HR_{regenerator,ref}$ = Reference regenerator heat requirement (3,763 kJ/kgCO_2)

$M_{CO_2,max}$ = Maximum CO_2 capacity of one absorber vessel (208.7 tonne/hr)

MEA_{makeup} = Flow rate of makeup sorbent (kg/hr)

$MEA_{makeup,ref}$= Reference flow rate of makeup sorbent (55.25 kg/hr)

The various indirect capital costs are then estimated as a function of the total PFC, while the preproduction costs are estimated as 1-month fixed and variable O&M costs (excluding electricity costs).

Table 12: Deep Carbon Capture Capital Cost Model Parameters and Nominal Values.

Capital Cost Element	**Unit**	**Value**[a]
General Facilities Capital (GFC)	% PFC	10
Engineering & Home Office Fees (E)	% PFC	7
Process Contingency Cost (C)	% PFC	10
Project Contingency Cost (F)	% (PFC+E+C)	20.0
Interest Charges (AFUDC)[b]		
Royalty Fees	% PFC	0.5
Preproduction (Startup) FOM Cost	1 month	1
Preproduction (Startup) VOM Cost	1 month	1
Miscellaneous Capital Cost (in Preproduction)	%TPI[c]	2
Inventory (Working) Capital	%TPC[c]	0.5
Financing Cost	%TPC	0
Other Owner's Cost	%TPC	0

a. Values are retrieved from IECM v 11.5.

b. $AFUDC = TPC * (\frac{(\frac{1+dis}{1+ei})^{N}-1}{N*(\frac{1+dis}{1+ei}-1)} - 1)$; *dis* is the discount rate, and *ei* is the inflation rate.

c. TPC is the total plant cost, and TPI is the total plant investment; TPC = PFC+GFC+E+C+F; TPI = TPC + AFUDC.

4.2. Operating and Maintenance Costs

The O&M Costs can be categorized into variable O&M costs and Fixed O&M costs. A structure of the O&M costs is presented in Figure 13. Variable O&M (VOM) costs account for the use of chemicals, water, and electricity for CO_2 system operation. Fixed O&M (FOM) costs account for the costs of labor, maintenance, and tax & insurance. The cost model parameters for VOM and FOM are shown in Table 13.

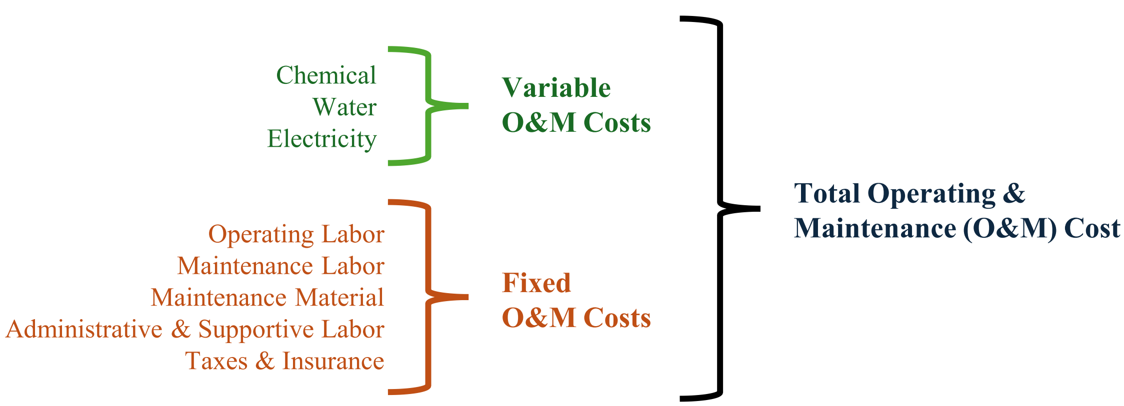


***Figure 13**: Structure of O&M costs of carbon capture system.*

Table 13: Parameters and Assumptions for Estimating O&M Costs.

O&M Cost Elements	Typical Value
Fixed O&M Costs	
Total Maintenance Cost	2.5% TPC
Maintenance Cost Allocated to Labor	40% of the total maintenance cost
Admin. & Support Labor Cost	30% of the total labor cost
Operating Labor Jobs per Shift	2 jobs/shift
Operable Labor Rate	$34.65/hr
Variable O&M Costs	
Reagent (MEA) Cost	$2,361/ tonne
Water Cost	$0.2721/ kliter
Activated Carbon	$2,205 /tonne
Caustic (NaOH)	$455 /tonne
Solid Waste Disposal Cost	$233/ tonne waste

a. Values are retrieved from IECM v 11.5.

The FOM costs include the costs of maintenance (materials and labor) and labor (operating labor, administrative, and support labor). These are estimated on an annual basis (M$/yr). The mathematical model for the fixed cost is given by Equations 36–38:

$$\mathrm{FOM} = \mathrm{FOM}_{\mathrm{labor,op}} + \mathrm{FOM}_{\mathrm{maint}} + \mathrm{FOM}_{\mathrm{admin}} \quad (36)$$

$$\mathrm{FOM}_{\mathrm{labor,op}} = \mathrm{C}_{\mathrm{labor,op}} * \mathrm{N}_{\mathrm{labor,op}} * 40(\mathrm{hrs/week}) * 52(\mathrm{weeks/yr}) \quad (37)$$

$$\mathrm{FOM}_{\mathrm{maint}} = \mathrm{TPC} * \mathrm{F}_{\mathrm{maint}}^{\mathrm{labor}} + \mathrm{TPC} * \mathrm{F}_{\mathrm{maint}}^{\mathrm{material}} \quad (38)$$

where,

FOM = Total fixed O&M cost, or annual fixed cost ($/yr)

$\mathrm{FOM}_{\mathrm{labor,op}}$ = Operating labor cost ($/yr)

$\mathrm{FOM}_{\mathrm{maint}}$ = Maintenance cost ($/yr)

$\mathrm{FOM}_{\mathrm{admin}}$ = Administrative & supportive cost ($/yr)

$\mathrm{C}_{\mathrm{labor,op}}$ = Operating labor rate ($/hr)

$\mathrm{N}_{\mathrm{labor,op}}$ = Number of operating labor (jobs/shift)

TPC = Total plant cost ($)

$\mathrm{F}_{\mathrm{maint}}^{\mathrm{labor}}$ = Fraction of maintenance cost allocated to maintenance labor ($/yr)

$\mathrm{F}_{\mathrm{maint}}^{\mathrm{material}}$ = Fraction of maintenance cost allocated to maintenance material ($/yr)

The VOM costs include the costs associated with chemicals, electricity, water, and waste disposal. These costs are estimated using Equations 39–42:

$$\mathrm{VOM}_{\mathrm{chem}} = \sum_{\mathrm{c=chemcials}} \mathrm{P_c} * \mathrm{M_c} * \mathrm{HPY} * \mathrm{CF}/10^6 \quad (39)$$

$$\mathrm{VOM_{elec}} = \mathrm{P_{elec}} * \mathrm{MW_{CC}} * \mathrm{HPY} * \mathrm{CF}/10^6 \quad (40)$$

$$\mathrm{VOM_{water}} = \mathrm{P_{water}} * \mathrm{M_{water}} * \mathrm{HPY} * \mathrm{CF}/10^6 \quad (41)$$

$$\mathrm{VOM_{wd}} = \mathrm{P_{wd}} * \mathrm{M_{wd}} * \mathrm{HPY} * \mathrm{CF}/10^6 \quad (42)$$

where,

$\mathrm{VOM_{chem}}$ = Chemical cost, which is the cost summation of sorbent, activated carbon, corrosion inhibitor, activated carbon, and caustic soda (M$/yr)

$\mathrm{VOM_{elec}}$ = Electricity cost (M$/yr)

$\mathrm{VOM_{water}}$ = Water cost (M$/yr)

$\mathrm{VOM_{wd}}$ = Reclaimer waste disposal cost (M$/yr)

$\mathrm{P_c}$ = Price of chemical, c ($/tonne)

$\mathrm{M_c}$ = Amount of chemical use, c (tonne/hr)

$\mathrm{P_{elec}}$ = Price of electricity ($/MWh)

$\mathrm{MW_{CC}}$ = Power use for carbon capture (MWe)

$\mathrm{P_{wd}}$ = price of waste disposal ($/tonne)

$\mathrm{M_{wd}}$ = amount of waste generated (tonne/hr)

$\mathrm{P_{water}}$ = price of water ($/kilter)

$\mathrm{M_{water}}$ = Process water requirement (tonne/hr)

Cost of MEA reagent: The makeup MEA requirement estimated in the performance model equals the difference between the total amount of sorbent loss and the amount of sorbent reclaimed. Total sorbent loss includes loss due to acid gas impurities, loss due to polymerization and oxidation, loss due to heat-stable salt formation, and loss in flue gas exhaust are calculating using Equations 43a–f.

$$MEA_{makeup} = \mathrm{Lost_{MEA}^{acid\ gas}} + \mathrm{Lost_{MEA}^{Polymerization}} + \mathrm{Lost_{MEA}^{Oxidation}} + \left(\mathrm{Lost_{MEA}^{HSS}} - \mathrm{Gain_{MEA}^{reclaimer}}\right) + \mathrm{M_{MEA}^{Fluegas\ Exhaust}} \quad (43a)$$

$$\mathrm{Lost_{MEA}^{acid\ gas}} = \left(2 * \mathrm{Remove_{SO_2}^{MEA}} * \eta_{SO_2}^{MEA} + 2 * \mathrm{Remove_{SO_3}^{MEA}} * \eta_{SO_3}^{MEA} + 2 * \mathrm{Remove_{NO_2}^{MEA}} * \eta_{NO_2}^{MEA} + \mathrm{Remove_{HCl}^{MEA}} * \eta_{HCl}^{MEA}\right) * \mathrm{MW_{MEA}}/1000 \quad (43b)$$

$$\mathrm{Lost_{MEA}^{Polymerization}} = 0.5 * \mathrm{LF_{MEA}} * \dot{m}_{CO_2}/1000 \quad (43c)$$

$$\mathrm{Lost_{MEA}^{HSS}} = \mathrm{Lost_{MEA}^{Oxidation}} * 1/1000 \quad (43d)$$

$$\mathrm{Lost_{MEA}^{Oxidation}} = 0.5 * \mathrm{LF_{MEA}} * \dot{m}_{CO_2}/1000 \quad (43e)$$

$$\mathrm{Gain_{MEA}^{reclaimer}} = 0.13 * \frac{\mathrm{MW_{MEA}}}{\mathrm{MW_{NaOH}}} * \dot{m}_{CO_2}/1000 \quad (43f)$$

where,

MEA_{makeup} = Mass flow rate of makeup MEA (tonne/hr)

$\text{Lost}_{\text{MEA}}^{\text{acid gas}}$ = MEA lost from flue gas acid (tonne/hr)

$\text{Lost}_{\text{MEA}}^{\text{Polymerization}}$ = MEA loss from polymerization (tonne/hr)

$\text{Lost}_{\text{MEA}}^{\text{HSS}}$ = MEA loss from the formation of heat-stable salts (tonne/hr)

$\text{Gain}_{\text{MEA}}^{\text{reclaimer}}$ = MEA gained from reclaimer regeneration (tonne/hr)

$\text{Lost}_{\text{MEA}}^{\text{Exahust}}$ = MEA lost in exhaust gas (tonne/hr)

$\text{Remove}_{\text{SO}_2/\text{SO}_3/\text{NO}_2/\text{HCl}}^{\text{MEA}}$ = Acid gas removed by MEA (kg-mole/hr)

$\eta_{\text{SO}_2/\text{SO}_3/\text{NO}_2/\text{HCl}}^{\text{MEA}}$ = Removal efficiency of acid gas by MEA (%)

MW_{MEA} = Molecular weight of MEA (g/mole)

MW_{NaOH} = Molecular weight of NaOH (g/mole)

$\dot{\text{m}}_{\text{CO}_2}$ = Mass flow rate of CO_2 product (tonne/hr)

1000 = Unit conversion from kg to tonne

$\text{Lost}_{\text{MEA}}^{\text{Oxidation}}$ = MEA lost in oxidation reactions (tonne/hr)

$\text{M}_{\text{Exhaust}}$ = Mass flow rate of exhaust gas (flue gas out) (tonne/hr)

$\text{F}_{\text{MEA}}^{\text{Exhaust}}$ = MEA loss in exhaust gas (ppm)

Cost of inhibitor: Addition of inhibitor makes it possible to use higher concentrations of MEA solvent in the system with minimal corrosion problems. Inhibitors are special compounds that come at a cost premium. The cost of the inhibitor is estimated as 20% of the cost of MEA as given by Equation 44.

$$\text{VOM}_{\text{inhibitor}} = \text{VOM}_{\text{sorbent}} * 0.2 \tag{44}$$

where,

$\text{VOM}_{\text{inhibitor}}$ = variable cost of inhibitor, (M$/Yr)

$\text{VOM}_{\text{sorbent}}$ = variable cost of MEA sorbent, (M$/Yr)

Cost **of activated carbon:** Activated carbon bed in the solvent circuit helps in the removal of long-chain/ cyclic polymeric compounds formed from the degenerated MEA. Typically, the average consumption amount is estimated to be about 0.075 kg C/tonne CO_2.

$$\text{M}_{\text{AC}} = \dot{m}_{CO_2} * 0.075/1000 \tag{45}$$

where,

M_{AC} = mass flow rate of activated carbon (tonne/hr)

0.075 = activated carbon use (kg C/tonne CO_2)

5. Case Study

Five pulverized coal case studies are included: (1) supercritical PC plant without biomass co-firing and a CO_2 capture system, (2) supercritical PC plant with a CO_2 capture system of 95% CO_2 capture rate, (3) supercritical PC plant with a CO_2 capture system of 99% CO_2 capture rate, (4) supercritical PC plant with 10% biomass co-firing and a CO_2 capture system of 95% CO_2 capture rate, and (5) supercritical PC plant with 10% biomass co-firing and a CO_2 capture system of 99% CO_2 capture rate. The IECM v13.0-beta is applied for power plant simulation. Illinois No. 6 and Switchgrass (SG) are the coal and biomass fuels selected for the Simulation. Furthermore, the simulation also specified the plant configuration, fuel as-delivered cost, etc. The following Table 14 includes input parameters and the major techno-economic results.

Several assumptions are made when modeling power plants. This study assumed that modeled plants are more than 30 years old, meaning the capital cost of the base plant and environmental control system is fully amortized. These modeled plants are modeled to have 30 years of remaining economic book life. The unit's capacity factor and primary heat input are assumed to remain unchanged after biomass co-firing and CCS retrofit. A retrofit multiplier of 1.15 is applied to the capital costs of newly installed CCS. All costs are expressed in constant 2018 dollars.

Table 14: Power Plant Configuration and Major Techno-Economic Inputs.

Parameter		**Value**				
Cases		PC Plant	PC Plant with 95% CCS	PC Plant with 99% CCS	PC Plant with 10% SG Co-firing and 95% CCS	PC Plant with 10% SG Co-firing and 99% CCS
Plant Location		U.S. Midwest Region				
Overall Plant Configuration	NOx Control System	Hot-Side SCR				
	Particulates Controls System	Fabric Filter				
	SO_2 Control System	Wet FGD				
	CO_2 Capture System	Not Applicable	Amine System			
	Cooling System	Wet Cooling Tower				
Fuel Properties	Coal Type	Illinois #6				
	Percentage of Coal in the Blend Fuel (wt.%)	100%	100%	100%	90%	90%
	Biomass Type	Not Applicable			Switchgrass	
	Percentage of Biomass in the Blend Fuel (wt.%)	0%	0%	0%	10%	10%
Fuel Cost Parameter	Coal Delivered Cost ($/tonne)	42.09				
	Bioimass Delivered Cost ($/tonne)	Not Applicable			112.8	

<table>
<tr><td>Overall Plant Performance Parameter</td><td>Capacity Factor (%)</td><td colspan="5">75</td></tr>
<tr><td rowspan="2">Overall Plant Financing Parameter</td><td>Costs Reported Year</td><td colspan="5">2018</td></tr>
<tr><td>Plant or Project Book Life (yr)</td><td colspan="5">30</td></tr>
<tr><td rowspan="6">Base Plant Performance Parameter</td><td>Gross Electrical Output (MW_g)</td><td>650</td><td>529.5</td><td>522.2</td><td>526.2</td><td>519</td></tr>
<tr><td>Primary Thermal Input (GJ/hr)</td><td colspan="5">5,744</td></tr>
<tr><td>Unit Type</td><td colspan="5">Supercritical</td></tr>
<tr><td>Steam Cycle Heat $Rate_{HHV}$ (kJ/kWh)</td><td>7,764</td><td>9,532</td><td>9,665</td><td>9,541</td><td>9,674</td></tr>
<tr><td>Boiler Firing Type</td><td colspan="5">Tangential</td></tr>
<tr><td>Boiler Efficiency (%)</td><td colspan="3">87.87</td><td colspan="2">87.41</td></tr>
<tr><td>Base Plant Capital Cost</td><td>Total Capital Requirement (TCR) Amortized (%)</td><td colspan="5" rowspan="5">100%</td></tr>
<tr><td>In-Furance NOx Control Capital Cost</td><td>TCR Amortized (%)</td></tr>
<tr><td>Hot-Side SCR NOx Control Capital Cost</td><td>TCR Amortized (%)</td></tr>
<tr><td>Particulates Controls Capital Cost</td><td>TCR Amortized (%)</td></tr>
<tr><td>SO_2 Control Capital Cost</td><td>TCR Amortized (%)</td></tr>
<tr><td rowspan="3">CO_2 Capture, Transport and Storage Configuration</td><td>CO_2 Capture Technology</td><td rowspan="9">Not Applicable</td><td colspan="4">Amine-based Deep Carbon Capture</td></tr>
<tr><td>Construction Time (yr)</td><td colspan="4">3</td></tr>
<tr><td>CO_2 Compression Stages</td><td colspan="4">8</td></tr>
<tr><td rowspan="6">CO_2 Capture, Transport and Storage Performance Parameter</td><td>Absorber CO_2 Removal Efficiency (%)</td><td>95%</td><td>99%</td><td>95%</td><td>99%</td></tr>
<tr><td>Lean CO_2 Loading (mole CO_2/mole sorb)</td><td colspan="4">0.2</td></tr>
<tr><td>Sorbent Losses (kg/tonne)</td><td colspan="4">0.4501</td></tr>
<tr><td>Total Absorber Intercooler Duty (MW_{th})</td><td>126.5</td><td>148.6</td><td>126.5</td><td>148.6</td></tr>
<tr><td>Liquid-to-Gas Ratio (mole/mole)</td><td>3.419</td><td>3.611</td><td>3.422</td><td>3.614</td></tr>
<tr><td>Gas Phase Pressure Drop (Mpa)</td><td>0.008382</td><td>0.008937</td><td>0.008366</td><td>0.008921</td></tr>
</table>

	Regenerator Heat Requirement (kJ/kg CO_2)		3,630	3,694	3,630	3,693
	Solvent Pumping Head (MPa)		0.2			
	Capture System Cooling Duty (tonne H_2O/tonne CO_2)		80.87	83.2	80.87	83.19
CO_2 Capture, Transport and Storage Cost Parameter	General Facilities Capital (%PFC[a])		10			
	Engineering & Home Office Fees (%PFC)		7			
	Process Contingency Cost (%PFC)		10			
	Project Contingency Cost (%(PFC+E+C)[b])		20			
	Royalty Fees (%PFC)		0.5			
	Inventory Capital (%TPC)		0.5			
	Capital Cost Process Area Retrofit Factor		1.15			
	CO_2 Transport Cost ($/tonne)		2.0–2.1			
	CO_2 Storage Cost ($/tonne)		3.0–3.1			

a. PFC is the process facilities capital.
b. E is the engineering & home office fees, and C is the process contingency cost.

The major results from case studies are shown below in Table 15. In Table 15, it is shown that the deployment of CCS for 95–99% CO_2 capture would decrease the net plant efficiency and net power output by 31–32%, when assuming constant thermal input. Meanwhile, CO_2 onsite emission intensity is reduced by 93 to 99% when a CCS is deployed for a 95 to 99% CO_2 capture rate. This significant emission mitigation comes with a high cost: total capital requirement of the plant increased by 118–126% and the LCOE increased by 190–203%, respectively.

Unlike the deployment of CCS, the additional deployment of 10% SG biomass co-firing modestly affects the plant's performance and cost. By incorporating the 10% co-firing, it further reduced the net power output and net plant efficiency by about 1% and increased the LCOE by about 5%. In addition, implementing 10% co-firing with the CCS system can further reduce the plant's greenhouse gas (GHG) emissions by about 35 to 46% on a life cycle basis. This co-firing retrofit increases the costs of CO_2 capture and avoided by about 8%, resulting in the ranges of $48.6 and $79.8–80.1 per tonne of CO_2, respectively.

Table 15: Summary and Comparison of Case Study Results.

Parameter	**Value**				
Cases	PC plant	PC Plant with 95% CCS	PC Plant with 99% CCS	PC Plant with 10% SG Co-firing and 95% CCS	PC Plant with 10% SG Co-firing and 99% CCS
Gross Power Output (MW_g)	650	529.5	522.2	526.2	519
Net Power Output (MW)	604.8	420	410.8	416.6	407.5
Net Plant Efficiency (%)	37.9	26.3	25.7	26.1	25.5
CO_2 Emission Intensity (kg/kWh)	0.83	0.059	0.012	0.059	0.012
Total Capital Requirement ($/kW$_{net}$)	2,062	4,495	4,664	4,522	4,692
Levelized Cost of Electricity ($/MWh)	29.7	86.1	89.9	90.7	94.7
CO_2 Avoided Cost ($/tonne)	Not Applicable	73.6	74.1	79.8	80.1
CO_2 Captured Cost ($/tonne)		44.8	44.9	48.6	48.6
Life Cycle Greenhouse Gas Emissions (kg CO_2eq/kWh)	0.91	0.20	0.16	0.13	0.087

Appendix A. Supplementary Data for Reduced-Order Performance Model

Appendix A presents information regarding the regression analysis summaries for Equations 4–19 in Section 3, including the L/G ratio, regeneration heat requirement, treated flue gas temperature, wash column outlet temperature, compressor and pump use, intercooler and wash column duties, MEA make-up, absorber pressure drop, cooling water requirement, and absorption and stripping process packing volumes, as shown in Tables A1–A16.

Table A1: L/G ratio Regression Analysis Summary Output (mole/mole).

Regression Statistics for the L/G Ratio								
Multiple R	0.9916							
R Square	0.9832							
Adjusted R Square	0.9828							
Standard Error	0.1430							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	4	208.5	52.12	2550	2.987E-153			
Residual	174	3.556	0.02044					
Total	178	212						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	-7.873	0.3814	-20.65	1.168E-48	-8.626	-7.121	-8.626	-7.121
CO_2 Content (mole%)	0.2808	0.002997	93.72	7.528E-151	0.2749	0.2868	0.2749	0.2868
Lean Loading (mole/mole)	14.46	0.4521	32.00	8.410E-75	13.57	15.36	13.57	15.36
CO_2 Removal Rate (%)	0.0479	0.003422	14.00	2.577E-30	0.04115	0.05466	0.04115	0.05466
Flue gas Inlet Temperature (°C)	0.0119	0.003820	3.110	2.185E-03	0.004342	0.01942	0.00434	0.01942

Table A2: Regression Analysis Summary Output for the Regeneration Heat Requirement (MW_{th}).

Regression Statistics for the Regeneration Heat Requirement								
Multiple R	0.9919							
R Square	0.9838							
Adjusted R Square	0.9835							
Standard Error	23.77							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	3	5,995,251	1998417	3536	2.656E-156			
Residual	175	98907	565.2					
Total	178	6094158						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	783.6	37.71	20.78	4.007E-49	709.2	858.0	709.2	858.0
Lean Loading (mole/mole)	-3951	188.1	-21.01	1.029E-49	-4322	-3580	-4322	-3580
CO_2 Flow Rate to Compressor (kmole/s)	-39.99	11.15	-3.586	4.350E-04	-62.00	-17.98	-62.00	-17.98
Lean solution Flow Rate (kmole/s)	6.623	0.3714	17.83	3.261E-41	5.890	7.356	5.890	7.356

Table A3: Regression Analysis Summary Output for the Absorber Outlet Temperature (°C) of Treated Flue Gas.

Regression Statistics for the Treated Flue Gas Temperature								
Multiple R	0.9492							
R Square	0.9010							
Adjusted R Square	0.8975							
Standard Error	1.925							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	6	5799	966.5	260.8	1.357E-83			
Residual	172	637.5	3.706					
Total	178	6436						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	41.65	2.436	17.10	6.091E-39	36.85	46.46	36.85	46.46
Flow rate of Flue Gas (tonne/hr)	-0.03634	0.002398	-15.16	1.676E-33	-0.04107	-0.03161	-0.04107	-0.03161
Captured CO_2 (tonne/hr)	0.04307	0.004013	10.73	6.732E-21	0.03515	0.05099	0.03515	0.05099
CO_2 Content (mol%)	2.949	0.2564	11.50	4.566E-23	2.443	3.455	2.443	3.4550
L/G (mole/mole)	-8.456	0.8702	-9.717	4.573E-18	-10.17	-6.74	-10.17	-6.74
Lean Loading (mole/mole)	60.15	12.37	4.863	2.590E-06	35.74	84.56	35.74	84.56
Flow Rate to Wash Column (tonne/hr)	0.03442	0.002276	15.12	2.102E-33	0.02993	0.03891	0.02993	0.03891

Table A4: Regression Analysis Summary Output for the Wash Column Outlet Temperature (°C).

Regression Statistics for the Wash Column Outlet Temperaure								
Multiple R	0.9914							
R Square	0.9828							
Adjusted R Square	0.9826							
Standard Error	0.5195							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	2	2720	1360	5040	4.367E-156			
Residual	176	47.50	0.2699					
Total	178	2768						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	43.32	1.489	29.100	3.647E-69	40.39	46.26	40.39	46.26
CO_2 Removal Rate (%)	-0.2983	0.01349	-22.11	1.103E-52	-0.3249	-0.2717	-0.3249	-0.2717
Absorber Outlet Temperature (°C)	0.5739	0.0070	81.70	6.968E-142	0.5600	0.5878	0.5600	0.5878

Table A5: Regression Analysis Summary Output for the Process Absorption Packing Volume (m^3).

Regression Statistics for the Process Absorption Packing Volume								
Multiple R	0.9848							
R Square	0.9699							
Adjusted R Square	0.9694							
Standard Error	401.1							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	3	906566643	302188881	1878	8.277E-133			
Residual	175	28153245	160876					
Total	178	934719888						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	-8991	921.9	-9.753	3.215E-18	-10811	-7172	-10811	-7172
CO_2 Removal Rate (%)	92.54	9.716	9.524	1.373E-17	73.36	111.7	73.4	111.7
Flow Rate of Flue Gas (kmole/s)	99.67	2.002	49.79	2.975E-105	95.72	103.6	95.72	103.6
Lean Solution Flow Rate (kmole/s)	27.73	0.9163	30.26	1.937E-71	25.92	29.54	25.92	29.54

Table A6: Regression Analysis Summary Output for the Process Stripping Packing Volume (m^3).

Regression Statistics for the Stripping Packing Volume								
Multiple R	0.9889							
R Square	0.9780							
Adjusted R Square	0.9778							
Standard Error	165							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	1	213853408	213853408	7858	1.367E-148			
Residual	177	4816896	27214					
Total	178	218670304						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	-411.5	33.09	-12.44	6.372E-26	-476.9	-346.2	-476.9	-346.2
Reboiler Duty (MWth)	5.924	0.06683	88.65	1.367E-148	5.792	6.056	5.792	6.056

Table A7: Regression Analysis Summary Output for the Cooling Water Requirement (tonne/hr).

Regression Statistics for the Cooling Water Requirement								
Multiple R	0.9871							
R Square	0.9744							
Adjusted R Square	0.9739							
Standard Error	2504							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	3	41738621645	13912873882	2218	5.977E-139			
Residual	175	1097639774	6272227					
Total	178	42836261418						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	85173	3974	21.43	8.239E-51	77329	93016	77329	93016
Lean Loading (mole/mole)	-432048	19822	-21.80	9.502E-52	-471169	-392927	-471169	-392927
Captured CO_2 (tonne/hr)	-82.79	7.59	-10.91	1.825E-21	-97.77	-67.81	-97.77	-67.81
Lean Solution Flow Rate (kmole/s)	863.85	39.13	22.07	1.867E-52	786.6	941.1	786.6	941.1

Table A8: Regression Analysis Summary Output for the MEA Make-up (kg/hr) (Flue Gas Exhaust).

Regression Statistics for the MEA Make-up								
Multiple R	0.9978							
R Square	0.9956							
Adjusted R Square	0.9956							
Standard Error	0.4810							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	2	9262	4631	20016	2.662E-208			
Residual	176	40.72	0.2314					
Total	178	9303						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	0.1788	0.1073	1.666	9.753E-02	-3.303E-02	3.906E-01	-3.303E-02	3.906E-01
Flow Rate of Flue Gas (tonne/hr)	4.510E-03	2.351E-05	191.86	2.494E-206	4.464E-03	4.557E-03	4.464E-03	4.557E-03
Lean Solution Flow Rate (tonne/hr)	-2.998E-04	1.239E-05	-24.20	7.110E-58	-3.243E-04	-2.753E-04	-3.243E-04	-2.753E-04

Table A9: Regression Analysis Summary Output for the Pumps Power Use (MW_e).

Regression Statistics the Solution Circulating Pumps								
Multiple R	0.9993							
R Square	0.9986							
Adjusted R Square	0.9986							
Standard Error	0.0165							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	1	34.58	34.58	127688	5.969E-255			
Residual	177	0.0479	0.0003					
Total	178	34.63						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	-0.01856	3.255E-03	-5.700	4.913E-08	-0.02498	-0.01213	-0.02498	-0.01213
Lean Solution Flow Rate (kmole/s)	0.01222	3.419E-05	357.3	5.969E-255	0.01215	0.01228	0.01215	0.01228

Table A10: Regression Analysis Summary Output for the Compressors Power Use (MW_e).

Multiple R	0.9975							
R Square	0.9950							
Adjusted R Square	0.9950							
Standard Error	1.1101							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	1	43785	43785	35529	6.410E-206			
Residual	177	218.1	1.232					
Total	178	44003						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	0.7991	0.2295	3.482	6.269E-04	0.3462	1.2520	0.3462	1.2520
Captured CO_2 (tonne/hr)	0.0885	0.0005	188.5	6.410E-206	0.08761	0.08946	0.08761	0.08946

Table A11: Regression Analysis Summary Output for the Wash Column Duty (MW_{th}).

Regression Statistics for the Wash Column Duty								
Multiple R	0.9561							
R Square	0.9142							
Adjusted R Square	0.9127							
Standard Error	13.70							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	3	349577	116526	621.2	5.089E-93			
Residual	175	32828	188					
Total	178	382405						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	-668.72	39.93	-16.75	3.459E-38	-747.5	-589.9	-747.5	-589.9
Flow Rate to Wash Column (tonne/hr)	0.01857	6.190E-04	30.01	6.672E-71	0.01735	0.01980	0.01735	0.01980
Absorber Outlet Temperature (°C)	6.598	0.1879	35.10	3.897E-81	6.227	6.969	6.227	6.969
CO_2 Removal Rate (%)	3.288	0.3574	9.200	1.060E-16	2.583	3.994	2.583	3.994

Table A12: Regression Analysis Summary Output for the Flow Rate to Wash Column (tonne/hr).

Regression Statistics for the Flow Rate to Wash Column								
Multiple R	0.9987							
R Square	0.9975							
Adjusted R Square	0.9974							
Standard Error	85.78							
Observations	179							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	4	503472620	125868155	17106	1.292E-224			
Residual	174	1280288	7358					
Total	178	504752908						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	2247	204.8	10.97	1.246E-21	1843	2651	1843	2651
Flow Rate of Flue Gas (tonne/hr)	0.9563	0.004183	228.6	1.243E-217	0.9481	0.9646	0.9481	0.9646
Lean Loading (mole/mole)	-1880	271.3	-6.930	7.889E-11	-2415	-1344	-2415	-1344
CO_2 Removal Rate (%)	-17.06	2.053	-8.309	2.640E-14	-21.11	-13.01	-21.11	-13.01
CO_2 Content (mole%)	-51.23	1.925	-26.61	3.169E-63	-55.03	-47.43	-55.03	-47.43

Table A13: Regression Analysis Summary Output for the Absorber Intercoolers Duty (MW_{th}) (PC cases).

Multiple R	0.95231							
R Square	0.90689							
Adjusted R Square	0.90214							
Standard Error	15.52							
Observations	104							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	5	230035	46007	190.9	7.034E-49			
Residual	98	23618	241					
Total	103	253654						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	622.7	70.65	8.814	4.510E-14	482.5	762.9	482.5	762.9
Absorber Outlet Temperature (°C)	-8.235	0.4108	-20.05	1.876E-36	-9.051	-7.420	-9.051	-7.420
CO_2 Content (mole%)	2.709	1.060	2.556	1.212E-02	0.6059	4.8116	0.6059	4.812
Lean Solution Flow Rate (tonne/hr)	-0.01380	0.001681	-8.214	8.853E-13	-0.01714	-0.01047	-0.01714	-0.01047
CO_2 Removal Rate (%)	-1.533	0.5915	-2.592	1.101E-02	-2.707	-0.3592	-2.707	-0.3592
Captured CO_2 (tonne/hr)	0.4335	0.03312	13.09	3.048E-23	0.3678	0.4992	0.3678	0.4992

Table A14: Regression Analysis Summary Output for the Absorber Intercoolers Duty (MW_{th}) (NGCC cases).

Regression Statistics Absorber Intercooler Duty (NGCC cases)								
Multiple R	0.9886							
R Square	0.9774							
Adjusted R Square	0.9758							
Standard Error	4.387							
Observations	75							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	5	57513	11502.5	597.6	2.457E-55			
Residual	69	1328	19.25					
Total	74	58841						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	-42.77	51.69	-0.8275	4.108E-01	-145.9	60.34	-145.9	60.34
Absorber Outlet Temperature (°C)	-1.238	0.9447	-1.311	1.942E-01	-3.123	0.6461	-3.123	0.6461
CO_2 Content (mol%)	3.353	1.360	2.465	1.618E-02	0.6398	6.065	0.6398	6.065
Lean Solution Flow Rate (tonne/hr)	0.03589	0.003909	9.181	1.397E-13	0.02809	0.04369	0.02809	0.04369
CO_2 Removal Rate (%)	0.9680	0.2239	4.322	5.082E-05	0.5212	1.415	0.5212	1.415
Captured CO_2 (tonne/hr)	-0.4586	0.06696	-6.849	2.492E-09	-0.5922	-0.3250	-0.5922	-0.3250

Table A15: Regression Analysis Summary Output for the Absorber Pressure Drop (kPa) (PC cases).

Regression Statistics for the Absorber Pressure Drop (PC Cases)								
Multiple R	0.9570							
R Square	0.9158							
Adjusted R Square	0.9124							
Standard Error	0.5467							
Observations	104							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	4	321.9	80.47	269.2	2.920E-52			
Residual	99	29.59	0.2989					
Total	103	351.5						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	1.502	0.3238	4.639	1.073E-05	0.8596	2.145	0.8596	2.145
Process Absorbtion Packing Volume (m^3)	8.722E-04	1.083E-04	8.057	1.808E-12	6.574E-04	1.087E-03	6.574E-04	1.087E-03
Lean Solution Flow Rate (tonne/hr)	-2.573E-04	3.802E-05	-6.767	9.355E-10	-3.327E-04	-1.818E-04	-3.327E-04	-1.818E-04
Flow Rate of Flue Gas (tonne/hr)	1.184E-03	1.835E-04	6.455	4.033E-09	8.204E-04	1.549E-03	8.204E-04	1.549E-03
Absorber Intercooler Duty (MWth)	9.929E-03	1.473E-03	6.742	1.052E-09	7.007E-03	1.285E-02	7.007E-03	1.285E-02

Table A16: Regression Analysis Summary Output for the Absorber Pressure drop (kPa) (NGCC cases).

Regression Statistics for the Absorber Pressure Drop (NGCC Cases)								
Multiple R	0.9998							
R Square	0.9995							
Adjusted R Square	0.9995							
Standard Error	0.1328							
Observations	75							
ANOVA								
	df	*SS*	*MS*	*F*	*Significance F*			
Regression	4	2682	670.5	38027	5.652E-116			
Residual	70	1.234	0.01763					
Total	74	2683						
	Coefficients	*Standard Error*	*t Stat*	*P-value*	*Lower 95%*	*Upper 95%*	*Lower 95.0%*	*Upper 95.0%*
Intercept	0.03554	0.03596	0.9882	3.264E-01	-0.03619	0.1073	-0.03619	0.1073
Process Absorbtion Packing Volume (m^3)	-2.027E-03	5.927E-05	-34.2020	2.002E-45	-0.00215	-1.909E-03	-2.145E-03	-1.909E-03
Lean Solution Flow Rate (tonne/hr)	4.187E-04	4.744E-05	8.8274	5.482E-13	0.00032	5.133E-04	3.241E-04	5.133E-04
Flow Rate of Flue Gas (tonne/hr)	4.980E-03	4.433E-05	112.3523	8.563E-81	0.00489	5.069E-03	4.892E-03	5.069E-03
Absorber Intercooler Duty (MWth)	-8.438E-03	1.659E-03	-5.0872	2.918E-06	-0.01175	-5.130E-03	-1.175E-02	-5.130E-03

Appendix B. Supplementary Information for Process Facilities Cost

Appendix B presents the data points and regression results of process facilities costs for amine-based carbon capture system in pulverized coal (PC) and natural gas combined cycle (NGCC) power plants, as a supplementary information to Equations 22–35 in Section 4.1.1. Figures B1–B11 illustrate process facility costs as functions of key performance parameters, with all data generated using IECM v11.5.

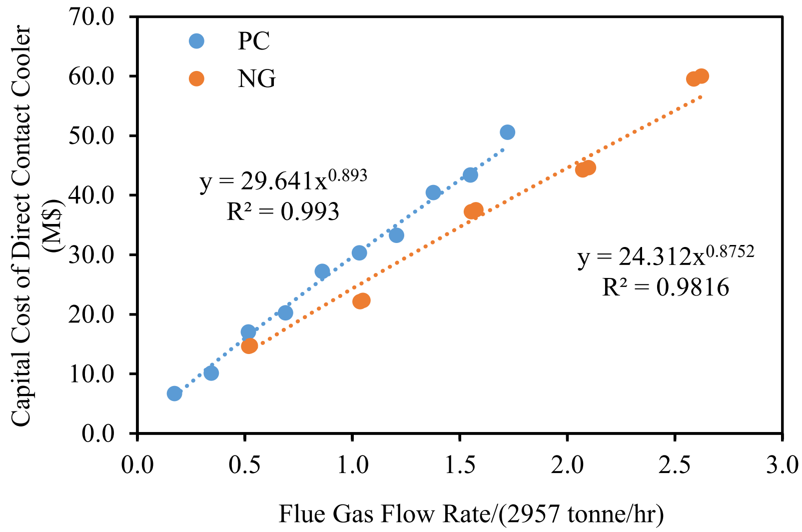


Figure B1: *Capital cost of the SO_2 polisher/direct contact cooler as a function of the flue gas flow rate.*

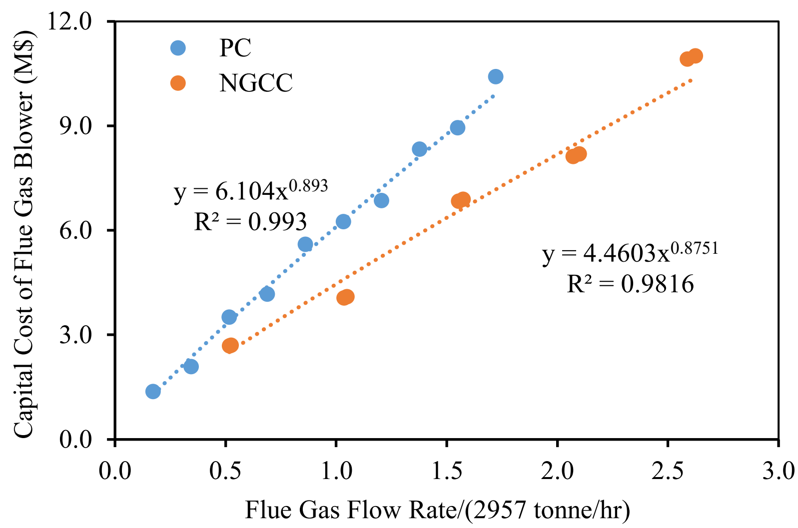


Figure B2: *Capital cost of a flue gas blower as a function of the flue gas flow rate.*

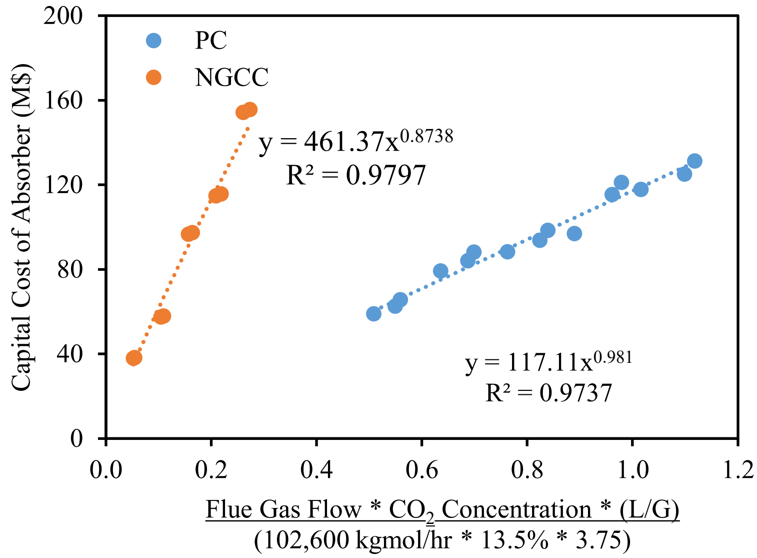


Figure B3: *Capital cost of absorber (vessels) as a function of the molar flow rate and* CO_2 *concentration of the inlet flue gas, and L/G ratio.*

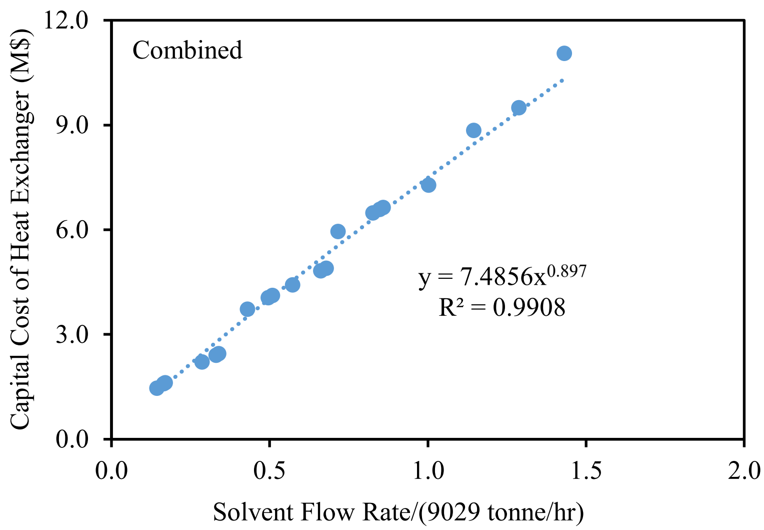


Figure B4: *Capital cost of rich/lean solvent cross heat exchanger as a function of the circulation flow rate of the solvent.*

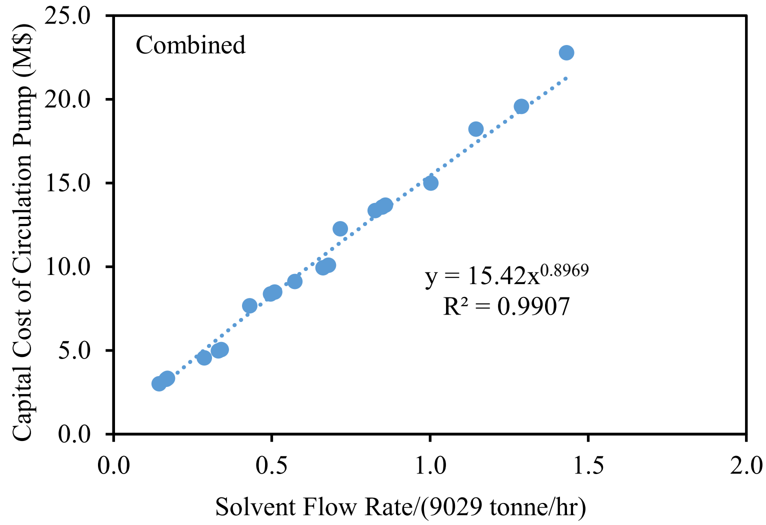


Figure B5: *Capital cost of circulation pumps as a function of the circulation flow rate of the solvent.*

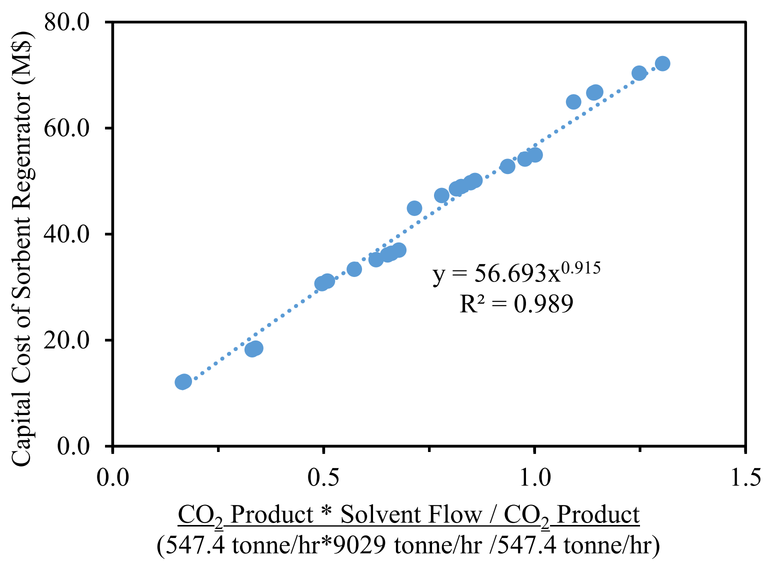


Figure B6: *Capital cost of sorbent regenerator as a function of the mass flow rate of CO_2 product and circulation flow rate of solvent.*

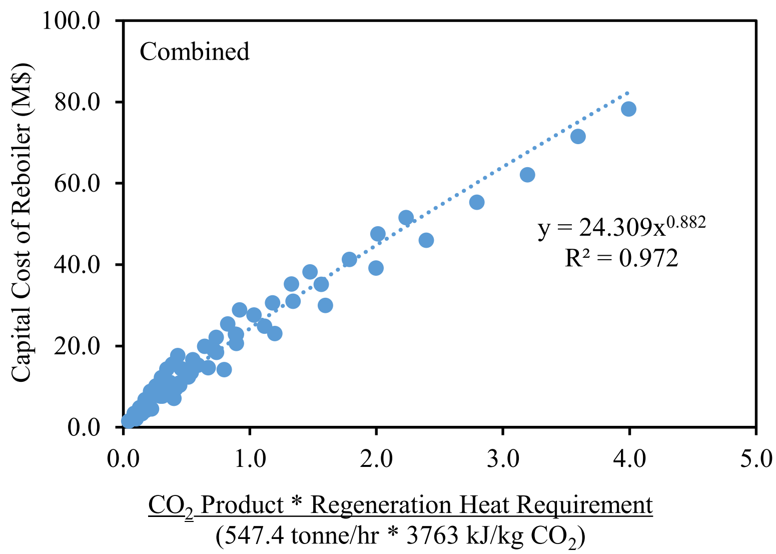


Figure B7: *Capital cost of reboiler (for regenerator) as a function of the mass flow rate of CO_2 product and regenerator heat requirement.*

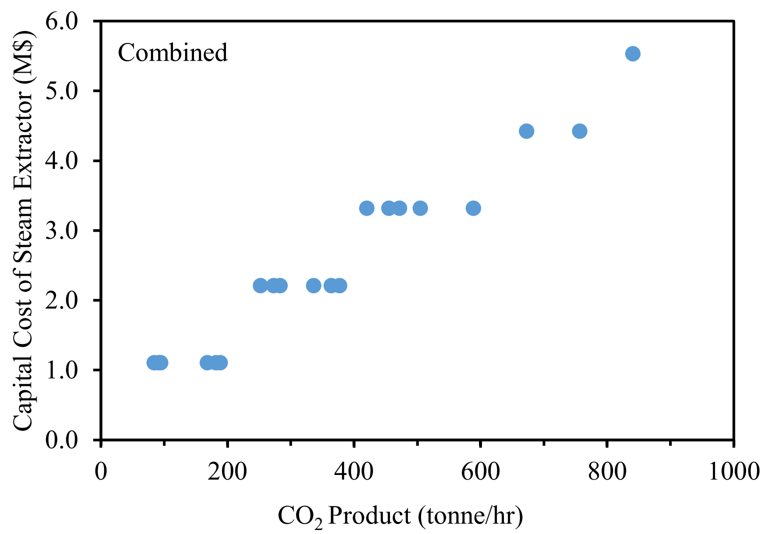


Figure B8: *Capital cost of steam extractor as a function of the mass flow rate of CO_2 product.*

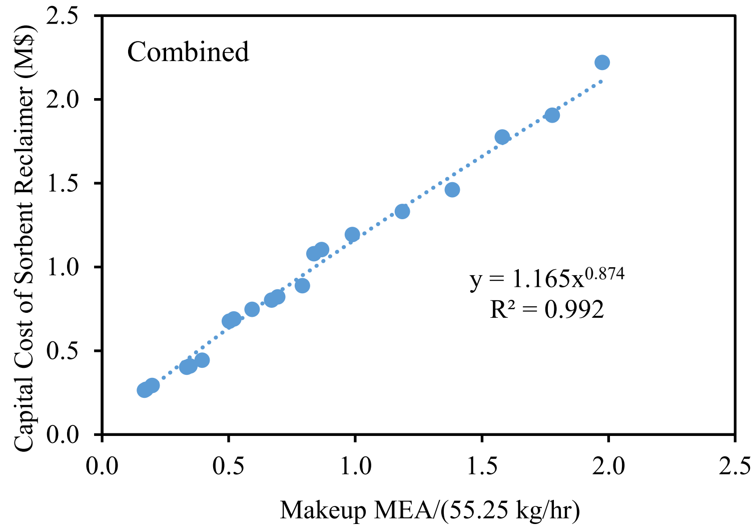


Figure B9: *Capital cost of sorbent reclaimer as a function of the mass flow rate of makeup MEA.*

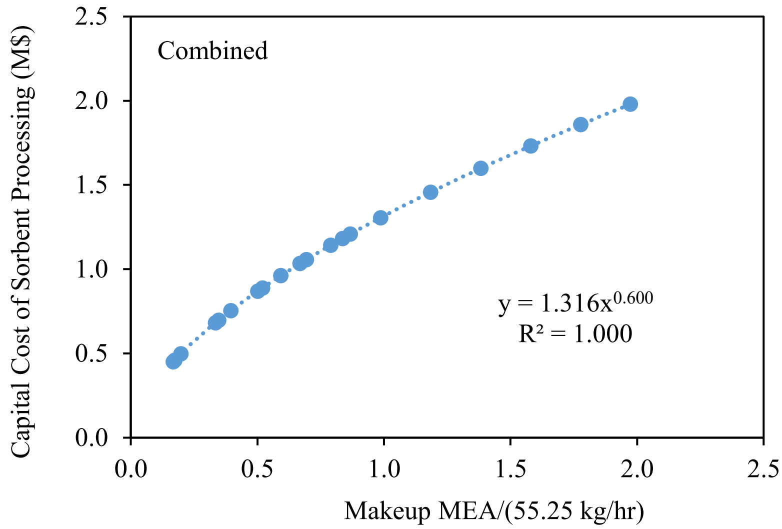


Figure B10: *Capital cost of Sorbent Processing Area as a function of the mass flow rate of CO_2 product.*

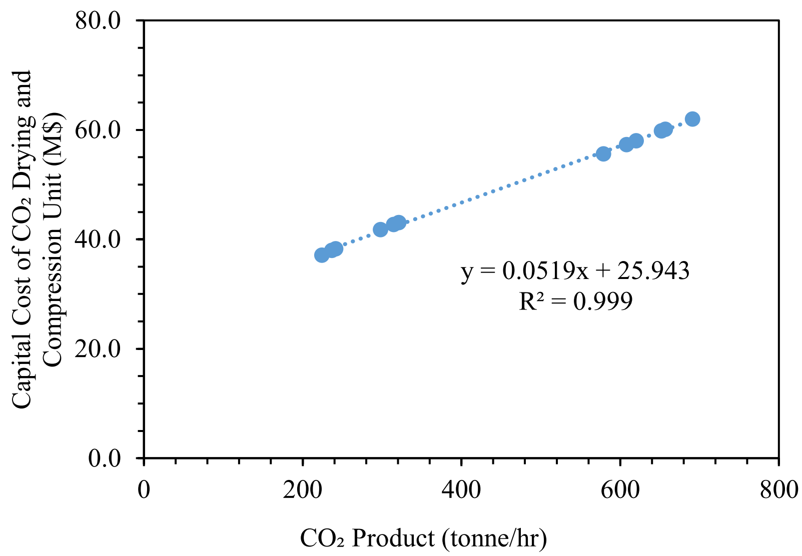


Figure B11: *Capital cost of CO_2 drying and compression unit as a function of the mass flow rate of CO_2 product.*

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