

Comprehensive thermo-economic optimization of a retrofit amine-based carbon capture process applied to a pulverized coal-fired power plant*

Alessio Tafone^{a†}, Benjamin Kanz^{ba}, Balkumar Basant Kumar Pillai^c, Jun Onn Khor^c, Tobias Massier^a, Harald Klein^b

a TUMCREATE, 1 CREATE Way, #10-02 CREATE Tower, 138602, Singapore

b Technical University of Munich, TUM School of Engineering and Design, Department of Energy and Process

c Energy Research Institute @ NTU, Nanyang Technological University, 1 Cleantech loop, 637141, Singapore

Accepted manuscript

[Comprehensive thermo-economic optimization of a retrofit amine-based carbon capture process applied to a pulverized coal-fired power plant](#) © 2026 by Alessio Tafone, Benjamin Kanz, Balkumar Basant Kumar Pillai, Jun Onn Khor, Tobias Massier and Harald Klein is licensed under the [CC BY- NC-ND 4.0](#) license. The final published version of this article is available at <https://doi.org/10.1016/j.energy.2026.142422>.

Abstract

Carbon dioxide abatement strategies are becoming crucial in the net-zero strategy defined by IEA, especially in those countries where the dependency on fossil fuels and specifically on coal is hard to abate in the short-medium term future. In this work, an in-house model, successfully developed and validated by the authors in UniSim®, was used to simulate an amine-based carbon capture system, designed to capture 90 % of the carbon dioxide emissions of a 500 MW_e supercritical coal fired power plant. Leveraging the interface between UniSim® and MATLAB®, a thermo-economic model has been developed and optimized to minimize energy consumption and the cost associated with the installation and the operation. The optimization is performed for a range of nominal lean CO₂ loading and absorber/stripper stages to investigate the influence of key design variables on the thermo-economic performance of the system. The retrofit case study is presented to demonstrate the usefulness of the proposed design and optimization framework and to provide practical guidelines and conceptual techno-economic insights for the design of a coal-fired power plant with CO₂ capture. The results show that the methodology is capable of capturing the optimal point of techno-economic design of the carbon capture plant with a specific reboiler heat duty of 3.26 MJ/kg_{CO2} and a LCOE of 107.5 USD/MWh_e, representing for the latter indicator a 64 % increase compared to the case without capture. The achieved reduction of the power plant carbon intensity (156 g_{CO2}/kWh_e) comes at the cost of an additional heat requirement that in the retrofit case study triggers a significant energy penalty of the power plant (18 %).

Keywords: Carbon capture, coal-fired power plants, MEA solvent, thermo-economic optimization, mitigation technologies.

Nomenclature

<i>Acronyms</i>		<i>SIC</i>	Specific investment cost [USD/kW _e]
CAPEX	Capital Cost [USD]	<i>Symbols</i>	
CC	Carbon capture	C	Cost [USD]

*The short version of the paper was presented at ICAE2024, Sep 1 – Sep 5, 2024. This paper is a substantial extension of the short version of the conference paper.

†Corresponding author. Tel.: (+65) 90150358. E-mail address: alessio.tafone@tum-create.edu.sg

CRF	Capital Recovery Factor	wt	Mass fraction [-]
FG	Flue Gas	C/M	Loading [$\text{mol}_{\text{CO}_2}/\text{mol}_{\text{MEA}}$]
HE	Heat Exchanger		
LCCO ₂	Levelized cost of CO ₂ [USD/ton _{CO₂}]	<i>Subscripts</i>	
LCOE	Levelized cost of energy [USD/MWh _e]	BM	Bare Module
MEA	Monoethanolamine	e	electric
OPEX	Operating cost [USD]	max	maximum
PCPP	Pulverized coal power plant	min	minimum
REC	Recuperator	reb	reboiler
SC	Specific consumption [$\text{MJ}_{\text{th}}/\text{kg}_{\text{CO}_2}$]	th	thermal

1. Introduction

Approaches to mitigate carbon dioxide emissions are playing an increasingly critical role in the net-zero roadmap outlined by the IEA [1]. This is especially true for countries where fossil fuel dependence, particularly on coal, remains deeply entrenched, making it difficult to transition to cleaner energy sources in the short to medium term. Indonesia has become one of the countries with the largest carbon dioxide (CO₂) emissions in the world because of its excessive dependence on coal for power generation, accounting for almost 50 % of the country's total CO₂ emissions [1]. Indonesia's primary energy consumption and power generation are quite similar to those in China and Australia, making it one of the largest per capita CO₂ emitters [1]. Indeed, China is responsible for nearly half of the global cumulative emissions from existing power plants, with the bulk of the remainder coming from other emerging economies, primarily due to their younger power plant fleets [2]. Most of the investments in these power generation assets in Asian countries were made in the past two decades, coinciding with periods of rapid economic expansion: coal plants are less than 20 years old on average, with China's fleet averaging just 13 years. By contrast, coal plants in Europe are about 35 years old on average, and in the United States, around 40 years [2].

To support the global effort needed to achieve carbon neutrality and thus limit global warming impact in the power generation sector, studies on CO₂ capture for coal-fired power plants represent a crucial alley to undertake for all those countries whose fuel mix heavily depends on coal and want to achieve significant decarbonization. According to IEA [2], in a sustainable development scenario where net carbon neutrality of the energy sector is achieved by 2070, the initial focus of carbon capture technology is on the retrofit of fossil fuel power plants accounting for approximately 40 % of the total CO₂ captured.

Nowadays, carbon capture technologies are at different maturity levels [3], especially depending on the application and the sector they are designed for. Chemical absorption is widely recognized as a practical method for retrofitting coal-fired power plants characterized by medium to low partial pressure of CO₂ in flue gas [4]. Nevertheless, although the removal of acid gases from process streams using amines is a well-established technology [5], flue gas scrubbing introduces numerous challenges that have yet to be sufficiently addressed for large-scale CO₂ abatement in the power generation sector [3]. Certainly, absorption methods for CO₂ capture need to advance to handle large volumes of flue gas efficiently, maintaining acceptable thermal efficiencies and keeping costs minimal. Aroonwilas and Veawab [6] technically analyzed the performance and the energy penalty induced by a post-combustion carbon capture unit applied to a 500 MW_e supercritical coal-fired power plant. Split flow configuration and a blend of MEA and MDEA were proposed as potential solutions to make the plant less energy intensive compared to the conventional MEA unit (36.8 % vs 43.9 % net electrical efficiency, respectively). Similarly, Idem et al.[7] experimentally studied the potential of mixed MEA/MDEA solvents for two different case studies mimicking two flue gas compositions: University of Regina pilot plant (natural gas combustion) and Boundary Dam CO₂ plant (coal combustion). In both cases, the novel MEA/MDEA mix ensured significant

reduction of the specific reboiler heat duty, especially for the coal combustion case study. Nevertheless, the chemical stability of the MEA/MDEA solvent has been shown to be heavily compromised compared to the single MEA solvent. Stec et al. [8] constructed and operated a post-combustion carbon capture plant at the Jaworzno II pulverized coal power plant in Poland. Conventional MEA solvent at 30 % wt concentration was employed to investigate the effect of split flow at the stripper inlet and inter-stage absorber intercooling over the specific reboiler heat duty. The experimental results showed that a 5 % reduction of the specific reboiler heat duty was achieved. Recently, Frimpong et al. [9] experimentally tested the post-combustion carbon capture concept with a pilot plant installed at the pulverized coal power plant unit at Kentucky Utilities E.W. Brown Power Generation Station in Harrodsburg, UK. The experimental runs were carried out to evaluate the specific reboiler heat duty and the CO₂ absorption capacity for different process conditions among which the mass flow rate of the working fluid, the stripper pressure and the inlet gas feed CO₂ concentration to the absorber. The measured specific reboiler heat duty ranged from 2.9 to 3.3 GJ/ton_{CO2} under different test conditions for a 30 % wt MEA concentration.

As also experimentally demonstrated above, current research and development of amine-based CO₂ absorption processes are mainly focused on improving the design of CO₂ capture processes with the aim of reducing the reboiler heat duty of the stripper, and consequently on the energy penalty in power plants [10][11]. These improvements reported in literature (e.g. absorber intercooling, stripper overhead compression split rich amine flow, lean amine flash etc.) claim to significantly reduce up to 20 % the energy demand. Still, in the vast majority of the previous studies, the economic aspect is neglected [12]. Indeed, the complexity of the process comes at a cost due to the upscaling or addition of new equipment that cannot be neglected. Despite presenting insightful outcomes regarding the process enhancement of the post-combustion CO₂ capture for a coal power plant, Zhao et al. [13] did not provide any insights regarding the heat source required for the regenerative process of the solvent. Furthermore, the economic feasibility of the proposed novel configurations was not considered. Similarly, Oh et al. [10] proposed a superstructure in order to comparatively assess four different configurations of the MEA-based post-combustion CO₂ capture plant for two different case studies (600 MW_e coal power plant and 435 MW_e combined cycle gas turbine) under partial load conditions. Confirming the results achieved by Aroonwilas and Veawab [6], flue gas splitting resulted in the most energy efficient solution; however, capital cost considerations were, once again, not included in the methodology.

Recently, Julio et al. [12] tried to overcome this research gap by proposing a techno-economic analysis of an amine-based carbon capture system. However, their work lacked to provide a consistent case study, as they did not address the origin of flue gas the carbon capture system needed to process and, above all, did not consider the origin of the heat source for the solvent regeneration in the stripper, as in Refs. [13] and [10].

Determining the origin of this heat source is a fundamental part of the design of an MEA absorption plant. The approach used here is to include auxiliary heat as part of the design to keep nearly unaltered the original electrical power output of the power plant. Neglecting the source of heat for the reboiler might trigger incorrect design and economic estimation of the plant since: i) the capital cost of the plant would be drastically underestimated without considering the cost of the auxiliary boiler; ii) overlooking the additional emissions from the auxiliary boiler erroneously leads to significant plant undersizing.

Different from the state-of-the-art mostly focusing on improved process design and new solvents, this paper deals with the optimal techno-economic design of an amine-based post-combustion carbon capture plant aiming to partially capture CO₂ emissions from a coal-fired power plant. The primary objective of the design methodology is to minimize the levelized cost of electricity (*LCOE*) of the power plant, enforce critical constraints to achieve a 90 % CO₂ capture rate and comply with the physical limits of the thermodynamic processes. In order to explore the effect of different design parameters heavily affecting energy consumption and plant cost, different process configurations and design parameters have been evaluated and techno-economically compared with a baseline case scenario representing the business-as-usual case study, namely the power plant without a carbon capture system.

2. Methodology

This study provides a comparative performance (techno-economic and environmental) analysis of the pulverized coal power plant system implementing an amine-based carbon capture process with a reference Rankine cycle system [14], given in this section. The following sub-sections discuss the Rankine cycle and absorption/desorption process models, design objectives, and critical constraints.

2.1. System configuration

A 500 MW_e supercritical Rankine cycle is taken as a reference to this plant. The plant is fed with Indonesian bituminous coal from South Sumatera, which was chosen because of its high calorific value among other coal samples from Indonesia [15]. The bituminous coal is air-dried before being combusted. The compositions of the South Sumatera coal are listed in **Table 1**. The compositions from the ultimate and proximate analysis were cross-checked on UniSim software and the difference in calorific value is less than 2.3 %.

Table 1. Composition of South Sumatera coal [15].

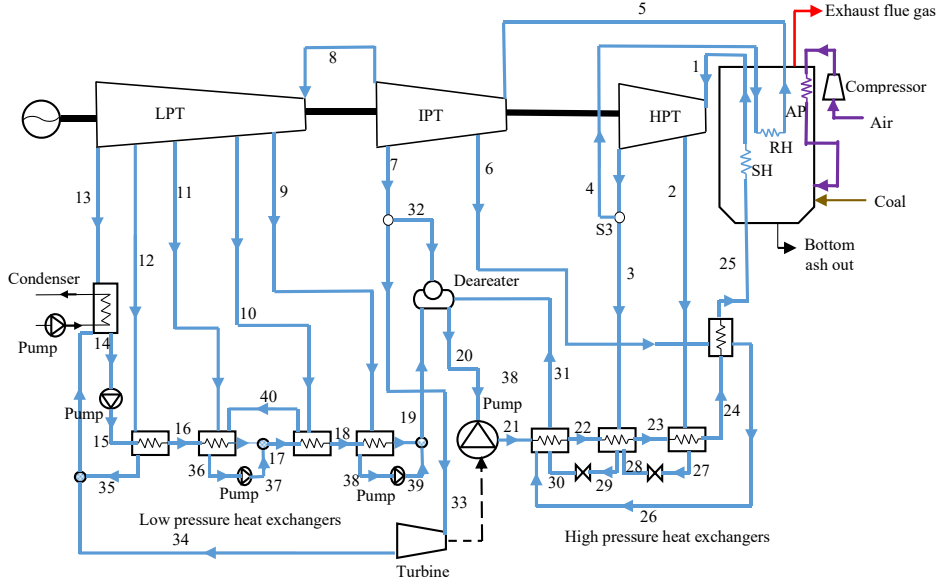
Component	Percentage
Proximate analysis	
Fixed carbon (adb)	43.42
Volatile matter (adb)	39.74
Ash (adb)	7.36
Moisture content (adb)	9.48
Ultimate analysis	
C (daf)	67.27
H (daf)	5.01
O (daf)	26.36
N (daf)	0.97
S (daf)	0.39
Gross calorific value [15]	25,330 kJ/kg
Gross calorific value (simulation)	25,900 kJ/kg

*adb = air-dried basis; daf = dry ash-free basis.

The schematics of the Rankine cycle of the coal-fired power plant, as shown in **Fig. 1**, was adapted from the work carried out by Reddy et al [16]. It mainly consists of a thermal boiler, a condenser with a centrifugal pump, a set of high-pressure feed pumps and a section of high, intermediate and low-pressure turbines (HPT, IPT and LPT) to produce electricity. In addition, a de-aerator and multiple high pressure and low-pressure feed water heaters were also included, along with other accessories to enhance the performance of power plant.

The schematic of the amine-based carbon capture process is reported in **Fig. 2**. The flue gas streams from the pulverized coal-fired power plant and the auxiliary boiler are fed to the absorber after being first cooled down and desulfurized (not shown in Fig.2). The CO₂ is absorbed in contact with the MEA solvent producing a CO₂ rich mixture. The latter exiting at the top of the absorber column is treated in a water wash section to recover the MEA/H₂O solution and avoid the release of the solvent into the environment. The aqueous phase, rich in CO₂, is pumped, passed through the recuperative heat exchanger (REC) and fed into the stripper column. Here the amine solvent is regenerated by providing heat via saturated steam (2.8 bar) at the bottom of the column (reboiler) that strips the CO₂ off from the processed stream. The lean amine solution at the bottom of the stripper column is then recirculated through the REC heat exchanger and recycled to the absorber where the CO₂ capture cycle starts again. In order to be effectively stored, the rich CO₂ stream at the top of the stripper column is cooled down in the condenser and then passed through the

compression station where the CO₂ is finally compressed to a final pressure of 110 bar by means of three intercooled stages.



Legend

	Fuel		Mixer
	Air		Splitter
	Flue gas		Heat exchanger
	H ₂ O stream		Pump
	Air compressor		Nozzle
	Steam turbine		Generator

Fig. 1. Schematic layout of the coal-fired power plant. Adapted from Ref. [16].

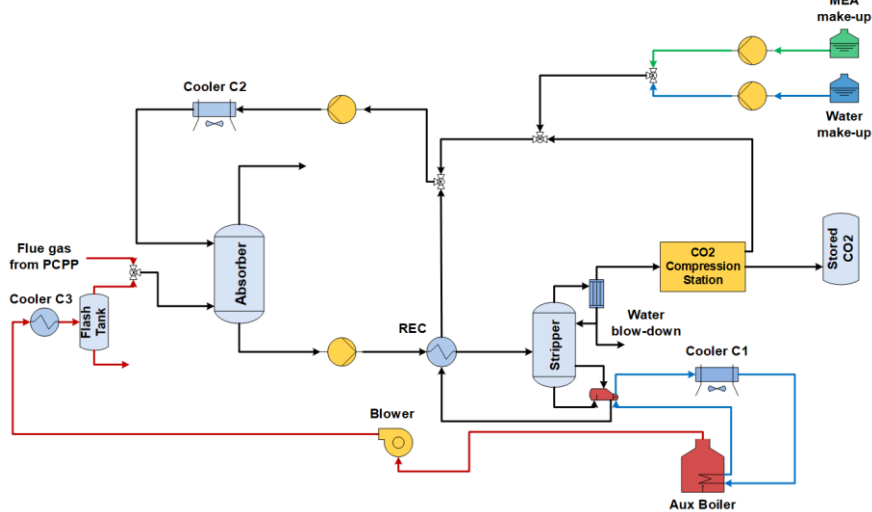


Fig. 2. Schematic layout of the amine-based carbon capture process.

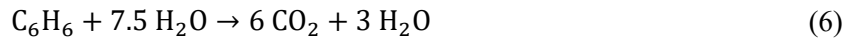
2.2. System modeling

2.2.1. Steam boiler and Rankine cycle

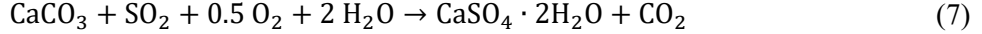
The power plant considered as the baseline case in our work is a supercritical pulverized coal-fired power plant (PCPP) with a nominal capacity of 500 MW_e. The power plant model built on UniSim consists of a coal combustion model, a Rankine cycle model and a post-combustion carbon capture model. The key operating conditions are reported in **Table 2**.

Table 2. Key operating conditions of the PCPP model implemented in Unisim [16].	
Components	Operating conditions
Air compressor	Outlet pressure: 1.04 bar
Combustor	Reactor condition: Adiabatic
	Operating pressure: 1.013 bar
	Calculation option: chemical and phase equilibrium
	No pressure drop
Steam turbine (HPT/IPT/LPT)	Cycle type: Supercritical
	Inlet pressure (HP/IP/LP): 300 bar/57.20 bar/8.361 bar
	Inlet temperature: 590 °C/610 °C/330.8 °C
	Isentropic efficiencies (HPT/IPT/LPT): 89 %
Condenser	Operating pressure: 0.05 bar
	Cooling water outlet temperature: 32.3 °C
	No pressure drop

The complete combustion of coal can be broken down and summarized with the following chemical equations [17]:



Sulfur dioxide is removed from the flue gas stream with a wet limestone desulfurization process, which is represented in the equation as follows:



The power plant's combustor and the auxiliary boiler models have been implemented to compute the flue gas temperature, pressure, mass flow rate and molar composition (**Table 3**), representing thus the boundary conditions of the amine-based absorption process.

Table 3. Exhaust gas molar composition.

Chemical formula	Value	Unit
H_2O	0.068	mol/mol
CO_2	0.154	mol/mol
O_2	0.033	mol/mol
N_2	0.745	mol/mol

2.2.2. Post-combustion carbon capture cycle

The post-combustion CO_2 capture thermodynamic cycle considered in this study is an absorption/desorption process using a chemical solvent MEA to remove CO_2 in the flue gases from the PC power plant. The capture plant using 29 wt% MEA is modeled in UniSim®, using the DBR Amines Pkg® with the equilibrium-based calculation using the Kent-Eisenberg model. This approach is suitable for gas-sweetening processes containing ethanolamines and has been proven to be effective for evaluating different amine process configurations [18]. The simulation parameters and techno-economic assumptions are delineated in **Table 4**, following an extensive review of the literature regarding systems operating with MEA [10,12,19,20].

Table 4. Technical specifications.

Parameters	Value	Unit
CO_2 removal rate	90	%
Flue gas mass flow rate	584.2	kg/s
Flue gas temperature	35	°C
Flue gas pressure	1.01	bar
MEA wt	29	%
Pump isentropic efficiency	75	%
Compressor polytropic efficiency	$f(PR)$	%
Absorber # stages	6-18	-
Absorber pressure	1	bar
Stripper # stages	6-18	-
Stripper pressure	2.8	bar
MEA Degradation Rate	0.15	$\text{kg}_{\text{MEA}}/\text{ton}_{\text{CO}_2}$
ΔT Pinch point REC	10	K
Column packing	Mellapak 250	-

In addition, the following assumptions have been considered valid throughout the whole investigation:

- SO_2 and NO_x components are completely removed from the flue gas stream.
- Coal combustion is complete in 20 % excess air.
- Water and MEA balances are achieved using a water/MEA make-up control system.
- Gravitational pressure drop is considered by introducing pressure valves at the inlet of both absorber and stripper.
- Pressure drops in heat exchangers are calculated as a function of the incoming pressure.
- The compressor polytropic efficiency is a function of the pressure ratio (PR) according to the correlation developed by Wilson [21], as specified in **Table 4**.

2.3. Thermo-economic analysis and optimization

A detailed thermo-economic analysis is performed to evaluate the economic penalty due to the installation and operation of the carbon capture (CC) plant. The assumptions, merits and terms used in this part are listed in **Table 5**. The total investment cost of the PCPP was calculated from the specific investment cost figures available in literature [1]. Conversely, since the amine-based absorption plant is not yet a commercially available technology for large flue gas flow rates, the total investment cost of the CC plant was computed by applying the module costing technique to calculate the bare module cost of each CC component (C_{BM}), modified by the chemical engineering plant cost index ($CEPCI$) to reflect inflation effects of current time from past year X :

$$CAPEX = \sum C_{BM,i} \left(\frac{CEPCI_{2022}}{CEPCI_X} \right) \quad (8)$$

Table 5. Techno-economic assumptions.

Economic Parameters	Value	Unit system	Reference
<i>SIC PCPP</i>	1200	USD/kW _e	[1]
<i>Coal price</i>	118.91	USD/ton	[22]
<i>MEA price</i>	1.22	USD/kg	[23]
<i>Water price</i>	1.66	USD/m ³	[24]
<i>Discount rate</i>	5	%	<i>Assumption</i>
<i>PCPP capacity factor</i>	80	%	[25]
<i>CEPCI 2022</i>	816	-	[26]
<i>Plant lifetime</i>	30	years	[25]

2.4. Cost correlations

To calculate the economic performance indicators, it is necessary to estimate the investment cost of the different post-combustion carbon capture plant components. To this end, the cost of each component is calculated based on specific size variables, such as the diameter and the height of the absorption and stripper column or the nominal power of the compressors. The total capital cost is the sum of the costs component and the working fluids cost (MEA and water).

Since such bare module component cost correlations are associated with significant uncertainties and the underlying data is usually not available and regionally specific, correlations from multiple sources are used for each component. Notably, in this paper four methods are used to calculate and compare the bare module

cost of each component, obtained from Turton et al. [27], Seider et al. [28], Couper et al. [29] and Peters et al. [30]. These cost correlations were chosen due to their capability to consider different variables affecting the final installation cost such as manufacturing materials, operating pressures and direct and indirect installation costs. The variability of the cost methods here is indicated by cost ranges with error bars in the plots, which correspond to the mathematical mean, as indicated. The use of multiple cost methodologies aims to reduce dependence on a single empirical correlation and to account for uncertainties associated with preliminary process cost estimation. The four equipment-cost correlations adopted in this study originate from different empirical datasets and therefore inevitably produce different absolute cost estimates. Since no universally accepted methodology exists for accurately predicting the capital cost of the present application, all four methods were considered to reduce dependence on any single empirical correlation. Equal weighting was adopted because no objective evidence is available to justify assigning greater confidence to one methodology over another. Accordingly, the arithmetic mean is employed as a representative estimate for the optimization, while the variability among the methods is retained to quantify the uncertainty associated with preliminary equipment costing. It should be noted that the proposed economic assessment is affected by uncertainties associated with component cost correlations. Although multiple established methodologies were employed to mitigate dependence on a single source, the resulting costs should be interpreted as preliminary estimates suitable for comparative techno-economic assessment rather than exact capital cost predictions. For the sake of comprehensiveness, a brief description of the cost correlations implemented for the main components (absorption/stripper columns, heat exchangers, compressors and vessels) is reported below.

2.4.1. Cost correlations from Turton et al. [27]

Based on the bare module factor technique methodology, Turton et al. [27] expresses the purchase base cost of the components (C_p^0 in USD), while operating at ambient pressure and carbon steel based, according to a logarithmic correlation:

$$\log C_p^0 = K_1 + K_2 \log(A) + K_3 [\log(A)]^2 \quad (9)$$

where A is the capacity or the size of the component under investigation (see Ref. [27] for the specific units) and K_1 , K_2 and K_3 are the correlation parameters derived from specific correlation charts in Ref. [27]. The values assumed for the correlation parameters are reported in **Table 6** for the main components.

Table 6. Correlation parameters obtained from Turton et al. for the main components [27].

Constant Parameters	K_1	K_2	K_3
<i>Compressors</i>	2.2897	1.3604	0.1027
<i>Storage vessels</i>	3.4974	0.4485	0.1074
<i>Columns</i>	3.4974	0.4485	0.1074
<i>Columns packing</i>	2.4493	0.9744	0.0055
<i>Pumps</i>	3.3892	0.0536	0.1538

Successively, in order to consider the manufacturing material, the operating pressure and the installation of the component, the bare module cost factor (F_{BM}) is applied to the purchase base cost of the components to determine the bare module cost (C_{BM}):

$$C_{BM} = F_{BM} C_p^0 \quad (10)$$

2.4.2. Cost correlations from Seider et al. [28]

Similar to Turton et al. [27], the cost correlation developed by Seider et al. [28] considers different parameters to determine the final installation cost of the component expressed in USD. Nevertheless, the methodology employed, and the factors used to determine the cost correlation significantly, differ from the one used by Turton et al. [27]. For instance, the total purchase cost (C_P) for packed columns is expressed as:

$$C_P = F_M C_V + C_{PL} + V_P C_{PK} + C_{DR} \quad (11)$$

Where F_M is the material factor and C_V is the purchase cost of the empty vessel expressed as a function of the weight of the shell and the two heads (W):

$$C_V = \exp\{K_1 + K_2[\ln(W)] + K_3[\ln(W)]^2\} \quad (12)$$

C_{PL} is the added cost for platforms and ladders calculated as a function of the column internal diameter (D_i) and the length of the shell (L):

$$C_{PL} = 300.9(D_i)^{0.63316} + (L)^{0.80161} \quad (13)$$

V_P , C_{PK} and C_{DR} are the volume and the installed cost of packing and the installed cost of liquid distributors to guarantee an adequate mixing in the packing, respectively.

Finally, as in Turton et al. [27], the bare module cost factor (F_{BM}) is applied to the total purchase base cost of the components to determine the bare module cost (C_{BM}):

$$C_{BM} = F_{BM} C_P \quad (14)$$

Similar procedures have been applied to calculate the cost of the other main components. The related methodologies here are omitted but a more exhaustive description can be found in [28].

2.4.3. Cost correlations from Couper et al. [29]

Belonging to the same component costing method group, the methodology proposed by Couper et al. [29] presents strong similarities with the one described by Seider et al. [28] but proposes different cost and correlation factors. Once again, for the sake of clarity, the total purchase cost (C_P in USD) methodology is here provided for the absorption/stripper columns:

$$C_P = 1.218[f_1 C_b + V_P C_P + C_{p1}] \quad (15)$$

where f_1 is the installation material factor and C_b is the base purchase cost of the empty vessel expressed as a function of the weight of the shell and the two heads (W):

$$C_b = \exp\{K_1 + K_2[\ln(W)] + K_3[\ln(W)]^2\} \quad (16)$$

V_P , C_P and C_{p1} are, as in Seider et al. [28], the volume and the installed cost of packing and the installed cost of liquid distributors, respectively.

As in Refs [27] and [28], the bare module cost factor (F_{BM}) is applied to the total purchased base cost of the components to determine the bare module cost (C_{BM}):

$$C_{BM} = F_{BM} C_P \quad (17)$$

2.4.4. Cost correlations from Peters et al. [30]

Differing from the references previously described, the purchased costs of the different components are provided by Peters et al. [30] in a graphical form. Therefore, empirical correlations have been extrapolated from those curves and reported in a numerical form in the techno-economic model. Consequently, a small additional uncertainty is introduced due to the digitization and fitting procedures. Nevertheless, this uncertainty is expected to be significantly smaller than the intrinsic uncertainty associated with preliminary equipment cost estimation correlations and therefore does not alter the overall conclusions of the present thermo-economic analysis.

For the absorption/stripper columns the total purchase cost C_P is reported as a function of the internal diameter (D_i) and the shell length (L), allowing subsequently to compute the bare module cost (C_{BM}):

$$C_P = [18038(D_i)^{1.1781}]L \quad (18)$$

$$C_{BM} = F_{BM}C_P \quad (19)$$

2.5. Key performance indicators

The main key performance indicators of the thermo-economic analyses are presented as follows. The first law analysis was implemented to compute the specific reboiler heat duty, computed as:

$$SC_{Reb} = \frac{\dot{Q}_{Reb}}{\dot{m}_{CO_2,stored}} \quad (20)$$

The levelized cost methodology has been applied to both the electricity and CO_2 commodities and describes the ratio of total annual cost of the investment in a power plant divided by its cumulative delivered electricity/stored CO_2 over the year:

$$LCOE = \frac{CAPEX_{CRF} + OPEX_{tot}}{El_{prod}} \quad (21)$$

$$\Delta LCOE = \frac{CAPEX_{CC}CRF + OPEX_{CC}}{El_{prod}} \quad (22)$$

$$LCCO_2 = \frac{CAPEX_{CC}CRF + OPEX_{CC}}{CO_{2,stored}} \quad (23)$$

where CRF [-] is the capital recovery factor, $\Delta LCOE$ [USD/kWh_e] is the delta $LCOE$ computed only taking into account the capital cost and the operating cost of the carbon capture plant, $OPEX_{tot}$ [USD/year] accounts for the operating and maintenance costs of the power plant and the CC plant ($OPEX_{O\&M,PCPP}$ and $OPEX_{O\&M,CC}$ [USD/year]), the operating cost related to the consumption of the coal for the steam boiler and the auxiliary boiler ($OPEX_{Coal,PCPP}$ and $OPEX_{Coal,CC}$ [USD/year]) and the operating costs related to the make-up of water and MEA ($OPEX_{H_2O,CC}$ and $OPEX_{MEA,CC}$ [USD/year]).

By varying the recirculation rate of the solvent through the lean CO_2 loading (C/M) entering the absorber (namely the ratio of the CO_2 molar flow rate to the MEA molar flow rate) and the number of stages of the absorption/stripper columns, different case studies are investigated. For each of those, an optimization routine has been developed in MATLAB[®] to identify the lean CO_2 loading ($\text{mol}_{\text{CO}_2}/\text{mol}_{\text{MEA}}$) and the number of absorption/stripper stages that minimize the $LCOE$. The optimization routine is based on the derivative-free optimization function ‘patternsearch’ available from the MATLAB[®] Optimization Toolbox. The step function and constraint tolerances for the current optimization problem were set to 10^{-4} . A detailed description of the ‘patternsearch’ algorithm is out of the scope of the present work but can be found in [31].

2.6. Optimization problem

In this section, the general structure of the optimization tool and the optimization technique is explained. The goal of optimization is to determine the optimal size of the post-combustion carbon capture plant components, which are defined as decision-making variables in Table 7. The optimal design and sizing of the carbon capture plant components are based on a three-level configuration (Fig. 3). The simulation level (UniSim), the patternsearch optimization level [32] and the interface for the communication between those two levels (MATLAB [33]). In particular, UniSim is a well-recognized process simulation software and platform that supports researchers and engineers to create steady-state and dynamic models for plant design, performance monitoring, troubleshooting, business planning, and asset management. Patternsearch is an optimization algorithm that minimizes a cost function (the objective function) without requiring a gradient by evaluating the outputs generated by an external simulation program, in the current case UniSim. Several optimization algorithms are available through MATLAB Optimization Toolbox and the user can choose the most adequate for the problem. Nevertheless, patternsearch has been chosen due to its capability to ensure a good compromise between the accuracy of the optimal solution and the computational time to achieve the optimum, as already previously assessed in literature [34]. Lastly, MATLAB acts as an interface between the UniSim process and the optimizer and streamlines the optimization process.

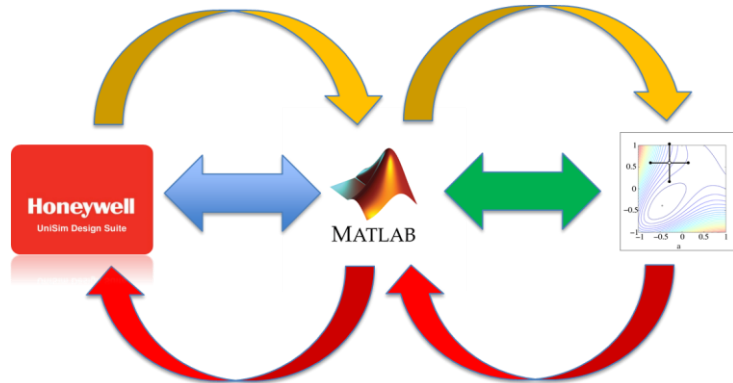


Fig. 3. UniSim-MATLAB-Patternsearch interfaces structure.

In the current work, the objective function to minimize has been set to be the levelized cost of energy as previously defined by eq. (21):

$$\text{Objective Function: Min } LCOE \quad (24)$$

The objective function is constrained to the energy and mass balance equation imposed in the UniSim model. It is worth underlining that potential revenues from the CO_2 utilization are not considered in the $LCOE$ and $\Delta LCOE$ equations as it is not within the scope of this work to merely maximize this profit by valorizing this commodity. Nevertheless, this component can be further analyzed in future work.

Table 7. Optimization search space.

Components	Search Space	
	Minimum	Maximum
<i>Absorber column</i>	6 stages	18 stages
<i>Stripper column</i>	6 stages	18 stages
<i>Lean C/M</i>	0.10	0.25

3. Results and discussion

This section discusses the optimal post-combustion carbon capture plant components capacities obtained and compares the economic and energy penalty over the base case. MATLAB minimizes the levelized cost of energy (i.e., the cost function of the optimization) through the patternsearch algorithm by performing UniSim simulations.

3.1. Models' validation

In order to guarantee an adequate level of confidence in the capacity of the models to accurately predict the systems performance and operation, the models were validated against data available in literature.

The combustion model is validated using the results from two different types of coal presented in the work of Suresh et al. [35]. In this work, coal is fully combusted before supplying heat to power the Rankine cycles. The coal combustion temperature of the UniSim model is validated against the published literature values [35] presented in **Table 8**.

Table 8. Validation of coal combustion temperature from numerical model against experimental data in [35].

	High-ash coal	Low-ash coal
Coal combustion temperature from Suresh et al. (°C)	1771.9	1921.4
Coal combustion temperature from UniSim model (°C)	1761.6	1921.8
Relative difference (%)	0.58	0.02

The Rankine cycle model was validated against the data provided by Reddy et al. [16]. The parameters represented in **Table 9** show that the streams of the simulated model in UniSim are in very good agreement with the model developed by Reddy et al. [16], presenting relative errors between -1.81 % and 1.15 %.

Table 9. Validation of the main streams parameters of the supercritical coal-fired power plant against data provided by Reddy et al. [16].

Stream ID	Temperature [K]			Pressure [bar]			Mass flowrate [kg/s]		
	Source	UniSim Value	Relative error (%)	Source	UniSim Value	Relative error (%)	Source	UniSim Value	Relative error (%)
1	863	863.15	-0.02	300	300	0	450	450	0
2	644.2	644.35	-0.02	76.8	76.82	-0.03	28.23	28.23	0
3	607.2	607.35	-0.02	58.37	58.36	0.02	33.37	33.37	0
4	607.2	607.35	-0.02	58.37	58.36	0.02	388.4	388.4	0

5	883	883.15	-0.02	57.2	57.2	0.00	388.4	388.4	0
6	769.4	769.55	-0.02	28.1	28.10	0.02	21.91	21.91	0
7	639.1	639.1	0	11.04	11.03	0.11	47.87	47.87	0
8	603.98	603.98	0	8.34	8.36	-0.26	318.62	318.62	0
9	529.9	529.9	0	4.42	4.43	-0.25	17.71	17.71	0
10	421.8	421.8	0	1.48	1.49	-0.54	17.72	17.72	0
11	352.4	352.43	-0.01	0.46	0.46	0	11.11	11.11	0
12	330.8	330.96	-0.05	0.18	0.18	0	13.89	13.89	0
13	305.9	306.03	-0.04	0.05	0.05	0	258.19	258.19	0
14	305.4	305.4	0.00	0.05	0.05	0	296.48	296.48	0
15	306	305.63	0.12	22	22	0	296.48	296.48	0
16	331	330.63	0.11	22	22	0	296.48	296.48	0
17	353	352.38	0.18	22	22	0	325.31	325.31	0
18	383	382.63	0.10	22	22	0	325.31	325.31	0
19	413.5	413.20	0.07	22	22	0	343.02	343.02	0
20	456	456.42	-0.09	10.82	10.82	0	450	450	0
21	462.5	462.5	0	344	344	0	450	450	0
22	489.05	489.05	0	344	344	0	450	450	0
23	524.92	524.92	0	344	344	0	450	450	0
24	550.54	550.54	0	344	344	0	450	450	0
25	554.85	554.85	0	344	344	0	450	450	0
26	576.8	585.02	-1.42	28.1	28.1	0	21.91	21.91	0
27	531.8	538.48	-1.26	75.26	75.26	0	28.23	28.23	0
28	531.8	538.41	-1.24	58.37	58.36	0.02	28.23	28.23	0
29	495.5	502.15	-1.34	57.2	57.2	0	61.59	61.6	-0.02
30	495.66	502.28	-1.34	28.1	28.1	0	61.59	61.6	-0.02
31	475.1	483.69	-1.81	27.54	27.54	0	83.5	83.51	-0.01
32	639.1	639.1	0	11.04	11.03	0.11	23.47	23.47	0
33	639.1	639.1	0	11.04	11.03	0.11	24.4	24.4	0
34	309.16	309.32	-0.05	0.06	0.06	0	24.4	24.4	0
35	330	330	0	0.18	0.18	0	13.89	13.89	0
36	351.6	351.6	0	0.46	0.46	0	28.82	28.83	-0.03
37	352	351.89	0.03	22	22	0	28.82	28.83	-0.03
38	412.8	412.8	0	4.42	4.42	0	17.71	17.71	0
39	413.14	413.12	0	22	22	0	17.71	17.71	0
40	381	376.62	1.15	1.46	1.46	0	17.72	17.72	0

Finally, the post-combustion amine-based carbon capture model was also validated against data provided by literature works adopting different CO₂ molar concentrations. Also in this case, the simulated results show good agreement with the data reported in literature (**Table 10**).

Table 10. Validation of the performance of the amine-based carbon capture UniSim model against literature data.

References	Reference value [MJ _{th} /kgCO ₂]	UniSim value [MJ _{th} /kgCO ₂]	Relative difference [%]
Oi [36]	3.59	3.61	-0.6
Lawal et al. [37]	3.48	3.42	1.7
Shirdel et al. [38]	3.65	3.56	2

3.2. Global techno-economic and environmental analysis

Table 11 lists the techno-economic parameters of the pulverized coal power plant with and without the post-combustion carbon capture system for the optimal techno-economic scenario. Since the CC plant requires extra thermal and electrical inputs for solvent regeneration and CO₂ compression, respectively, the electrical efficiency $\eta_{e,PP}$ drops by $\sim 18\%$, which is in good agreement with the results presented in literature [39]. Indeed, this reflects the high parasitic load caused by the total stored CO₂ compression phase. Notably, the total CO₂ emissions significantly increase due to the higher coal consumption related to the supplementary heating for solvent regeneration. However, most of this additional CO₂ is captured and prevented from reaching the atmosphere. The capital cost increases by $\sim 45\%$ with the addition of the carbon capture plant, reflecting the costs of the absorber, stripper, compressor, and other associated infrastructure, with a related *SIC* increase of $\sim 62\%$.

The higher investment costs and higher operating cost due to the solvent regeneration and CO₂ compression significantly increases the *LCOE* of the coal power plant by $\sim 62\%$ with a related leveled cost of CO₂ of 24.5 USD. As for the energy penalty, the *LCCO₂* value is consistent with findings presented in the literature (15-74 USD/tonCO₂ [12,19,40]).

Nevertheless, the carbon capture system is capable of reducing the carbon intensity of the plant by 84 % (156 vs 950 gCO₂/kWh_e) with an annual avoided CO₂ emission of 4409 ktonCO₂, making the coal power plant much cleaner and aligned with decarbonization targets. Moreover, the stored CO₂ boasts exceptional purity (99.9 %), making it highly suitable for a wide range of potential applications, including industrial processes, chemical synthesis, and sustainable product development.

Table 11. Techno-economic results of the PCPP w/ and w/o CC.

	Parameters	PCPP	PCPP&CC	Unit
Technical	<i>C/M</i>	-	0.24	-
	<i># of absorption stages</i>	-	12	-
	<i># of stripper stages</i>	-	18	-
	<i>Electric Power Output</i>	500.0	449.3	MW _e
	<i>Thermal Energy Input</i>	1180.8	1869.7	MW _{th}
	$\eta_{e,PP}$	42.34	24.03	%
Economical	<i>CAPEX</i>	600.0	871.6	MUSD
	<i>SIC</i>	1200.0	1939.8	USD/kW _e
	<i>LCOE</i>	65.7	107.5	USD/MWh _e
	<i>LCCO₂</i>	-	24.5	USD/tonCO ₂
Environmental	<i>CO₂ Mass flow rate</i>	132.0	194.2	kg/s
	<i>CO₂ Captured</i>	-	174.7	kg/s
	<i>CO₂ Purity wt</i>	-	99.9	%
	<i>CO₂ annually stored</i>	-	4408.7	ktonCO ₂ /year
	<i>CO₂ intensity</i>	950.3	156.2	gCO ₂ /kWh _e

The capital cost and the delta levelized cost of energy breakdown of the optimized configuration of the carbon capture plant are reported in **Fig. 4** and **Fig. 5**, respectively. Together, the stripper and the absorber account for over half (54.4 %) of the total capital cost. This is consistent with their central roles in the CO₂ capture process. The compression station also makes up a significant share of the cost. After CO₂ is captured, it must be compressed to high pressures for transportation (e.g. through pipelines) and storage (e.g., in geological formations). Furthermore, due to the power-matching retrofit strategy adopted in this work, the auxiliary boiler also significantly contributes to the final installation cost (16.6 %) due to the supplementary heating necessary to regenerate the solvent. This outcome is also confirmed by the distribution of *LCOE* components which shows the predominancy of the operating cost due to the supplementary heating accounting for almost 75 % (*OPEX_{Coal,CC}*) of the total cost components. Indeed, the relatively small size of the *CAPEX* component suggests that it contributes less to the total cost compared to *OPEX*, reflecting the high operating cost of carbon capture systems. This highlights the energy-intensive nature of amine-based capture systems as additional energy is needed for solvent regeneration, compression and other processes.

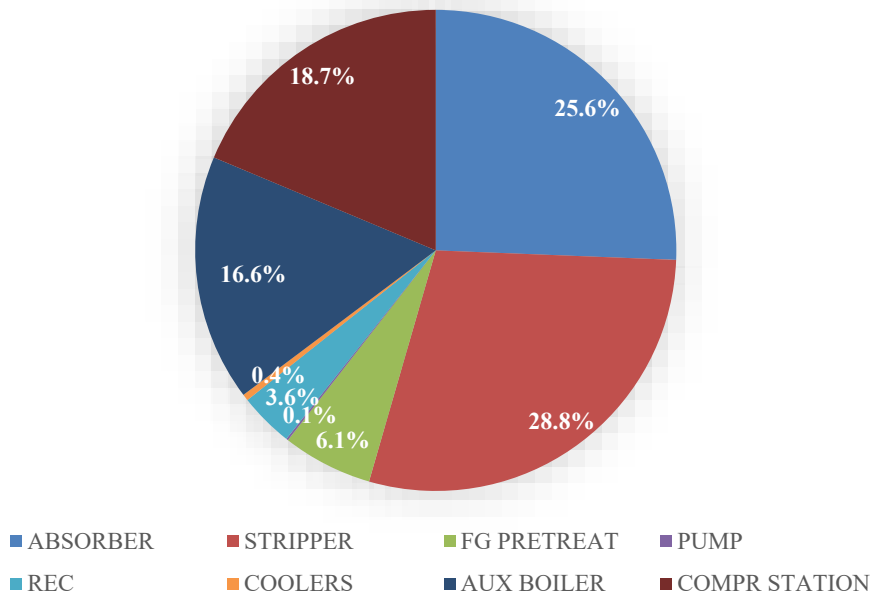


Fig. 4. CC Capital cost breakdown of the optimal configuration.

Table 12. CC Capital cost breakdown for the different CC components of the optimal configuration.

CC Components	Value [MUSD]
Stripper	78.36
Absorber	69.61
Compression Station	50.76
Auxiliary Boiler	45.00
FG Pretreatment	16.46
Recuperator	9.91
Coolers	1.14
Pump	0.35

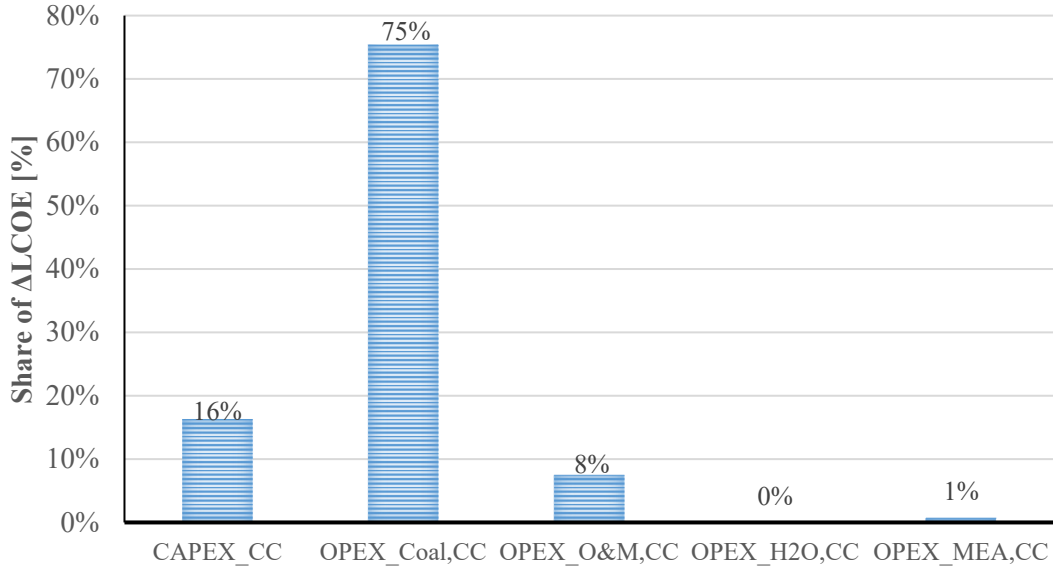


Fig. 5. Breakdown of the ΔLCOE by component of the optimal configuration.

3.3. Parametric analysis on C/M lean and absorption/stripper number of stages

As previously mentioned, the thermo-economic analysis is first conducted by carrying out a sensitivity assessment to optimize the levelized cost of energy. At a given flue gas composition and mass flow rate, the following three parameters are varied: 1) the lean CO_2 loading (C/M) entering the absorber, namely the ratio of the CO_2 molar flow rate to the MEA molar flow rate; 2) the number of absorption column stages; 3) the number of stripper column stages.

Fig. 6 shows the effect of the lean CO_2 loading on the specific reboiler heat duty for different numbers of absorption stages (e.g. $A_8 = 8$ absorption stages) at a fixed number of stripper stages (e.g. $S_{10} = 10$ stripper stages). The results show that by increasing the C/M value, which corresponds to a decrease in the MEA/ H_2O solution flow rate, the same fraction of CO_2 can be captured with a lower stripper duty until a minimum is achieved. Furthermore, the figure shows that the C/M value that minimizes the specific heat reboiler duty proportionally varies with the number of absorption stages: from $C/M = 0.15$ for A_8S_{10} scenario to $C/M = 0.22$ for $A_{16}S_{10}$ scenario.

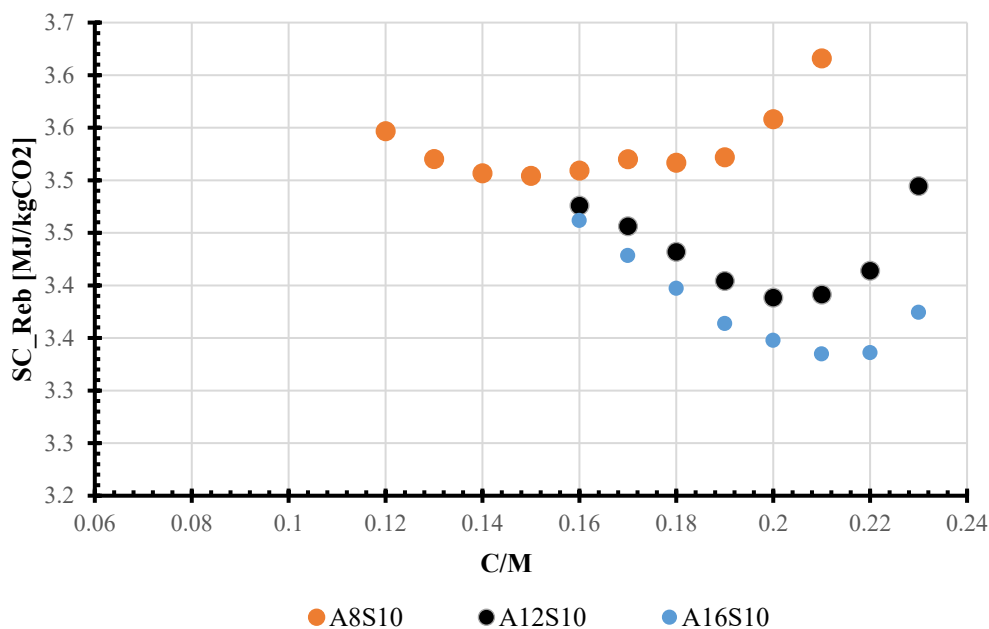


Fig. 6. Specific reboiler heat duty as a function of the lean CO₂ loading.

For each scenario where the number of stages has been varied, the minima of the specific reboiler duty are calculated and reported in the heat map of **Fig. 7**. As a general observation, absorption stages have a larger influence on the specific reboiler heat duty as compared to the stripper stages. Indeed, fixing $N_{strip} = 10$ and increasing N_{abs} from 6 to 18 reduces SC_{reb} from 3.7 MJ_{th}/kgCO₂ to 3.31 MJ_{th}/kgCO₂ with a 11 % decrease. Conversely, fixing $N_{abs} = 10$ and increasing N_{strip} from 6 to 18 produces a more modest decrease of the specific reboiler heat duty (~5 % from 3.48 MJ_{th}/kgCO₂ to 3.33 MJ_{th}/kgCO₂).

Further focusing the analysis on the absorption stages, **Fig. 8** provides a more direct visualization of their influence on the energy penalty induced by the carbon capture plant. As expected, higher absorption columns demand less solvent solution by approaching the minimum liquid requirement. Consequently, less MEA/H₂O solution flows to the stripping column, lowering the heat demand for the solvent regeneration. As above mentioned, an average 11 % decrease of the specific reboiler heat duty is achieved by increasing the number of absorption stages from 6 to 18. Nevertheless, the incremental decrease of energy consumption reduces stages by stages reaching an average minimum decrease of 0.3 % from 16 to 18 absorption stages. The results therefore suggest that increasing the number of absorption stages beyond a certain threshold offers diminishing returns, which may not justify the additional capital and operating costs.

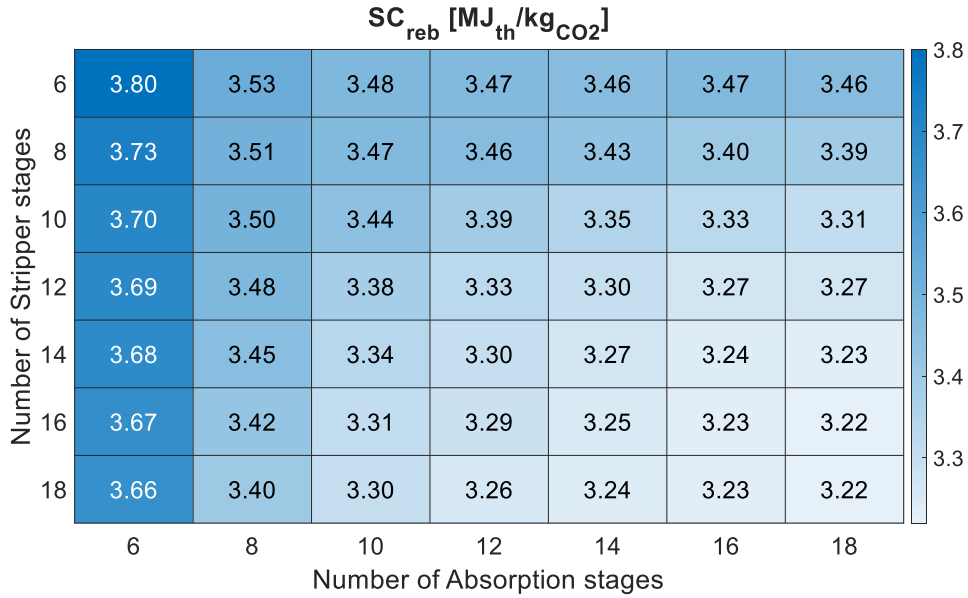


Fig. 7. Optimal specific reboiler heat duty for different number of absorption/stripper stages.

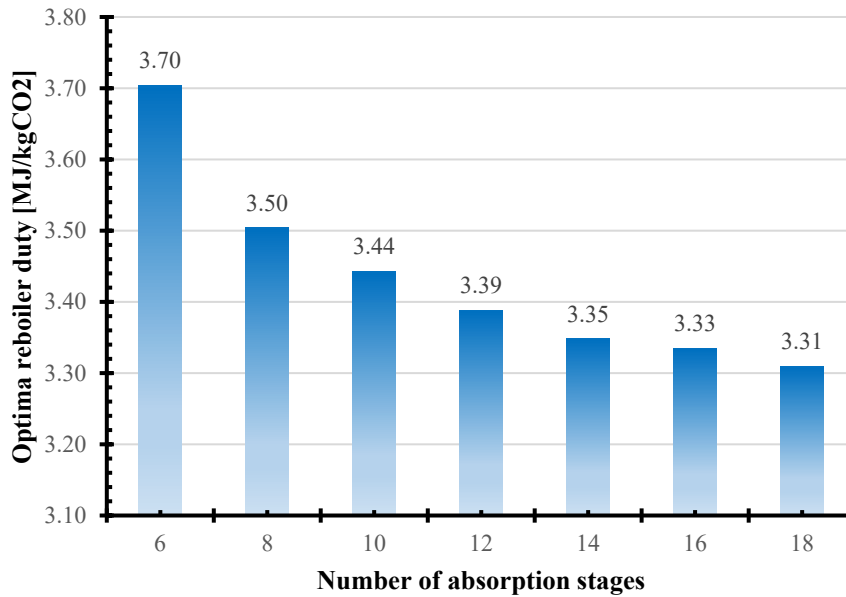


Fig. 8. Optimal specific reboiler heat duty for different number of absorption stages and fixed number of stripper stages ($N_{strip} = 10$).

This tendency is clearly reflected in the economics of the plant which in this work represents the main driver for the optimal design of the carbon capture plant. The minimization of the specific reboiler heat duty and the levelized cost of electricity are two competing goals, which implies that reducing the energy consumption at the reboiler triggers an extra investment cost (in this case related to the upscaling of the column height). Indeed, reporting the minima of the *LCOE* for the different case studies, **Fig. 9** represents these contrasting trends taking into account both the operating costs and the annualized capital cost. In

particular, the heat map illustrates the $LCOE$ as a function of the number of absorption stages and stripper stages. The $LCOE$ values range from 107.47 USD/MWh_e (the global minimum) to 115.84 USD/MWh_e. As mentioned in Section 2.4, the error bars reflect the above-mentioned variability of the four different capital cost methods.

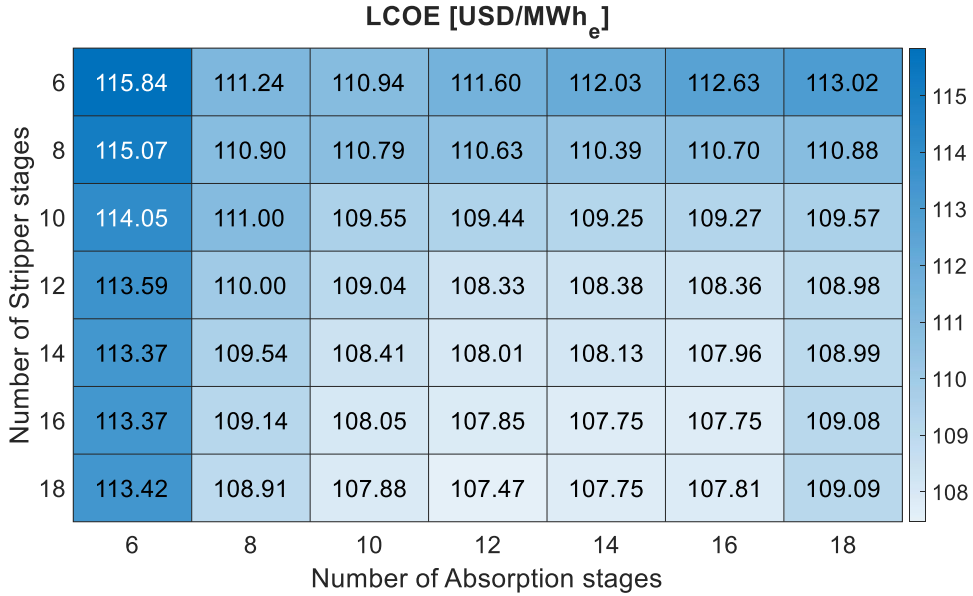


Fig. 9. Optimal levelized cost of energy for different number of absorption/stripper stages.

The figure shows that the relationship between the decision variables and $LCOE$ is significantly not linear: the increment of the absorber column height is beneficial until a minimum is achieved. After this turning point (usually located at around $N_{abs} = 12$ for any fixed number of stripper stages), the $LCOE$ starts to slightly increase, as shown in **Fig. 10** where the $LCOE$ values are reported for different numbers of absorption stages at fixed $N_{strip} = 10$.

A closer examination of the $LCOE$ breakdown specifically related to the carbon capture plant ($\Delta LCOE$) and its main contributing factors provides valuable insights into the observed phenomena. **Fig. 11** illustrates that at higher numbers of absorption stages (N_{abs}), the operating cost reductions, primarily driven by the improved specific reboiler heat duty, are completely counterbalanced by the corresponding increases in the annualized $CAPEX$ (where $CAPEX_{CC,ANN} = CAPEX_{CC,CRF}$). This trend becomes particularly evident starting from the scenario with $N_{abs} = 12$, where the capital cost begins to dominate as the primary driver of the overall cost. Beyond this point, the incremental reduction in heat duty requirements and associated operating costs fail to compensate for the rising capital expenses, leading to an overall increase in the $LCOE$. Consequently, after $N_{abs} = 12$, additional absorption stages result in diminishing returns, as the marginal efficiency gains are outweighed by the disproportionate growth in capital costs, thereby making the system less economically favorable.

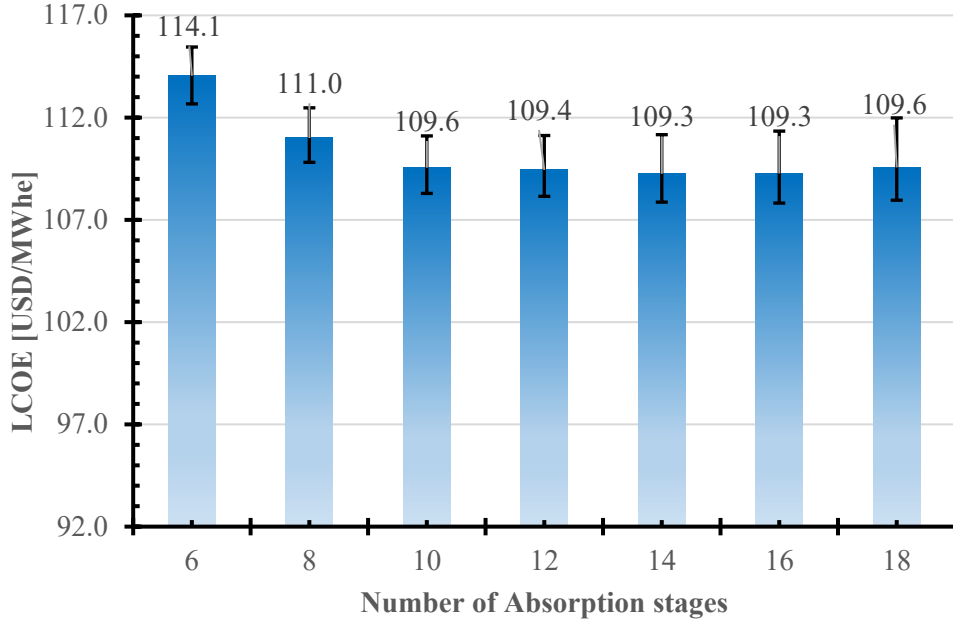


Fig. 10. Optimal levelized cost of energy for different number of absorption stages and fixed number of stripper stages ($N_{strip} = 10$).

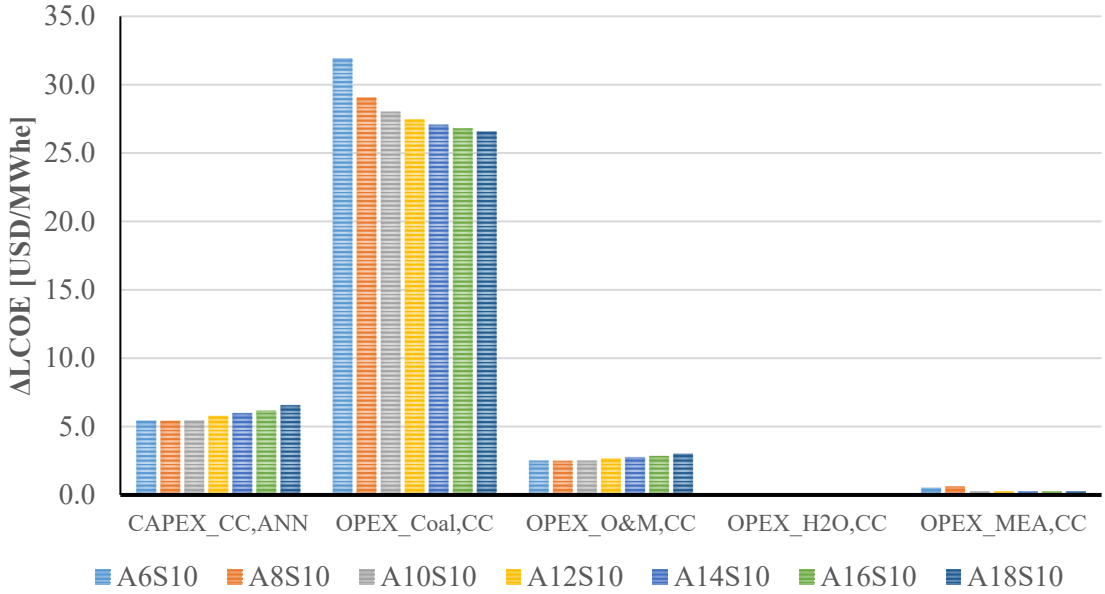


Fig. 11. $\Delta LCOE$ cost components distribution.

A similar trend can also be observed for the levelized cost of CO_2 , whose values are reported in the heat map of **Fig. 12**, once again as a function of the number of absorption/stripper stages. The $LCCO_2$ values range from 24.50 USD/ton $_{CO_2}$ to 27.40 USD/ton $_{CO_2}$. The minimum is observed at $N_{abs} = 12$ and $N_{strip} = 18$:

this matches the optimal configuration identified in the *LCOE* heatmap, reinforcing that this is a cost-effective balance for both electricity production and CO₂ capture. However, while the *LCOE* increases noticeably for a number of absorption stages higher than 12, primarily due to the rising capital costs associated with additional absorption stages, the *LCCO₂* shows a much less significant increase. This indicates that the cost of capturing CO₂ per ton remains relatively stable, even as the electricity generation cost rises. The stability in *LCCO₂* suggests that the marginal improvements in CO₂ capture efficiency and purity, achieved by adding more absorption stages, effectively offset the impact of higher capital and operating expenditures associated with these configurations. As a result, while the economic burden on electricity consumers grows due to higher *LCOE*, the specific cost of reducing CO₂ emissions, an essential metric for environmental and policy considerations, remains fairly consistent, demonstrating the resilience of the capture system in managing CO₂-related costs despite increasing system complexity.

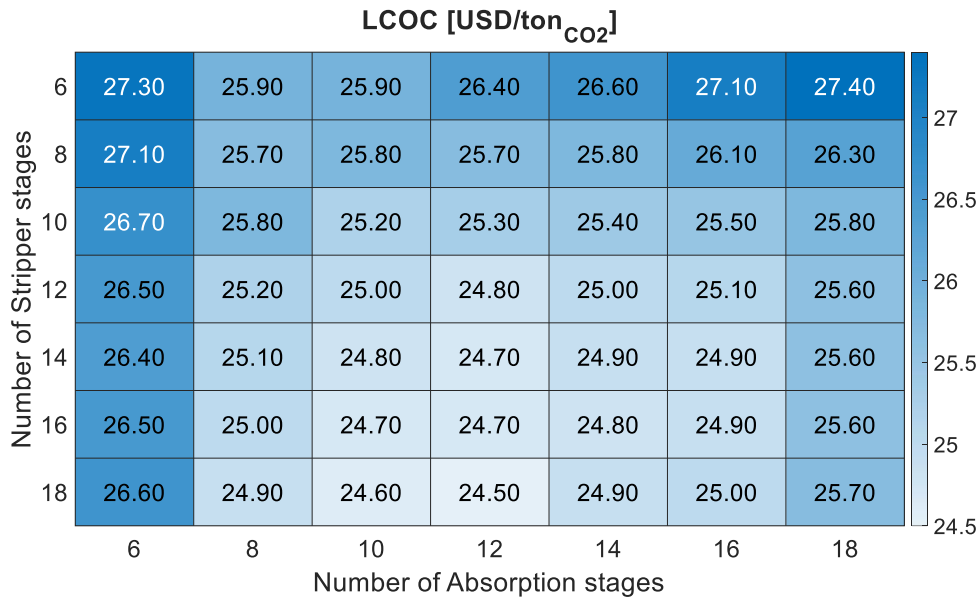


Fig. 12. Levelized cost of CO₂ for different number of absorption/stripper stages.

4. Conclusions

This paper investigated the thermodynamic and economic aspects of a pulverized coal power plant with post-combustion CO₂ absorption using MEA as a retrofit case study. To this end, an auxiliary boiler has been modeled to provide the additional thermal power needed for the solvent regeneration. By means of a comprehensive and detailed thermo-economic optimization, an optimal size of the absorption column has been identified that minimizes the levelized cost of energy (107.5 USD/MWh_e). It has been shown that the minimum of the specific reboiler heat duty does not necessarily guarantee a minimum *LCOE*. It is therefore of primary importance to establish a comprehensive model to ascertain the techno-economic performance of the CC plant. Globally, a net pulverized coal power plant efficiency of 24 % is achieved, about 18 % points less than the reference power plant without CO₂ capture, with a drop of the carbon intensity by 84 % (156 vs 950 g_{CO2}/kWh_e). In conclusion, this work provides a comprehensive design and decision-support framework for post-combustion CO₂ capture retrofits, enabling the identification of economically and energetically optimal configurations while explicitly accounting for the source of regeneration heat and its associated costs and emissions. To this end, it facilitates capturing the required trade-off between the need

for reducing both the energy consumption of the carbon capture plant and capital/operating costs, main goals for post-combustion capture systems.

Beyond the specific case study investigated, the main strength of the proposed methodology lies in its comprehensive thermo-economic framework, which simultaneously considers thermodynamic performance, economic feasibility, and environmental impact during the design phase. In contrast to many previous studies focusing primarily on reducing the reboiler heat duty, the present work explicitly accounts for the origin of the regeneration heat by integrating an auxiliary boiler into the system design. This allows a more realistic estimation of the capital costs, operating costs, additional CO₂ emissions, and overall plant performance. Furthermore, the results demonstrate that the configuration minimizing the specific reboiler heat duty does not necessarily correspond to the minimum levelized cost of electricity, highlighting the importance of performing integrated thermo-economic optimization when designing post-combustion carbon capture systems. The proposed framework therefore provides a practical and robust decision-support tool for the retrofit and optimization of large-scale carbon capture applications.

Despite the valuable insights provided by this study, some limitations should be acknowledged. The analysis was performed for a specific retrofit case consisting of a 500 MW_e supercritical coal-fired power plant using MEA as the solvent, and therefore the quantitative results may differ for other plant sizes, operating conditions, or solvent technologies. In addition, the economic assessment relies on established cost correlations that are suitable for preliminary techno-economic evaluations but remain subject to uncertainty. Furthermore, the optimization framework considers an auxiliary boiler as the sole source of regeneration heat and focuses on a limited set of design variables. Future work will extend the optimization framework by simultaneously considering additional design and operating variables, together with multi-objective thermo-economic-environmental optimization and alternative regeneration heat supply technologies (e.g. high temperature heat pump) to significantly reduce both the operating cost linked to supplementary heating and the related CO₂ extra production.

Acknowledgements

This work was financially supported by the Singapore National Research Foundation under its Campus for Research Excellence and Technological Enterprise (CREATE) programme (Grant No.: NRF2022-ITC004-001).

References

- [1] World Energy Outlook 2023 – Analysis - IEA n.d. <https://www.iea.org/reports/world-energy-outlook-2023> (accessed May 29, 2024).
- [2] IEA. Energy Technology Perspectives 2020 - Special Report on Carbon Capture Utilisation and Storage CCUS in clean energy transitions 2020. https://www.oecd.org/en/publications/energy-technology-perspectives-2020-special-report-on-carbon-capture-utilisation-and-storage_208b66f4-en.html (accessed March 18, 2025).
- [3] Hong WY. A techno-economic review on carbon capture, utilisation and storage systems for achieving a net-zero CO₂ emissions future. *Carbon Capture Science & Technology* 2022;3:100044. <https://doi.org/10.1016/J.CCST.2022.100044>.
- [4] Wilberforce T, Baroutaji A, Soudan B, Al-Alami AH, Olabi AG. Outlook of carbon capture technology and challenges. *Science of The Total Environment* 2019;657:56–72. <https://doi.org/10.1016/J.SCITOTENV.2018.11.424>.
- [5] Zhu W, Ye H, Zou X, Yang Y, Dong H. Analysis and optimization for chemical absorption of H₂S/CO₂ system: Applied in a multiple gas feeds sweetening process. *Sep Purif Technol* 2021;276:119301. <https://doi.org/10.1016/J.SEPPUR.2021.119301>.
- [6] Aroonwilas A, Veawab A. Integration of CO₂ capture unit using single- and blended-amines into supercritical coal-fired power plants: Implications for emission and energy management. *International*

- Journal of Greenhouse Gas Control 2007;1:143–50. [https://doi.org/10.1016/S1750-5836\(07\)00011-4](https://doi.org/10.1016/S1750-5836(07)00011-4).
- [7] Idem R, Wilson M, Tontiwachwuthikul P, Chakma A, Veawab A, Aroonwilas A, et al. Pilot Plant Studies of the CO₂ Capture Performance of Aqueous MEA and Mixed MEA/MDEA Solvents at the University of Regina CO₂ Capture Technology Development Plant and the Boundary Dam CO₂ Capture Demonstration Plant. *Ind Eng Chem Res* 2005;45:2414–20. <https://doi.org/10.1021/IE050569E>.
- [8] Stec M, Tatarczuk A, Więclaw-Solny L, Krótki A, Ciężko M, Tokarski S. Pilot plant results for advanced CO₂ capture process using amine scrubbing at the Jaworzno II Power Plant in Poland. *Fuel* 2015;151:50–6. <https://doi.org/10.1016/J.FUEL.2015.01.014>.
- [9] Frimpong RA, Nikolic H, Bahr D, Kiran G, Liu K. Pilot scale testing of an advanced solvent in a 0.7 MWe post-combustion CO₂ capture unit. *International Journal of Greenhouse Gas Control* 2021;106:103290. <https://doi.org/10.1016/J.IJGGC.2021.103290>.
- [10] Oh SY, Binns M, Cho H, Kim JK. Energy minimization of MEA-based CO₂ capture process. *Appl Energy* 2016;169:353–62. <https://doi.org/10.1016/J.APENERGY.2016.02.046>.
- [11] Ahn H, Luberti M, Liu Z, Brandani S. Process Simulation of Aqueous MEA Plants for Post-combustion Capture from Coal-fired Power Plants. *Energy Procedia* 2013;37:1523–31. <https://doi.org/10.1016/J.EGYPRO.2013.06.028>.
- [12] Julio AAV, Castro-Amoedo R, Maréchal F, González AM, Escobar Palacio JC. Exergy and economic analysis of the trade-off for design of post-combustion CO₂ capture plant by chemical absorption with MEA. *Energy* 2023;280:128004. <https://doi.org/10.1016/J.ENERGY.2023.128004>.
- [13] Zhao B, Liu F, Cui Z, Liu C, Yue H, Tang S, et al. Enhancing the energetic efficiency of MDEA/PZ-based CO₂ capture technology for a 650 MW power plant: Process improvement. *Appl Energy* 2017;185:362–75. <https://doi.org/10.1016/J.APENERGY.2016.11.009>.
- [14] Garcia DA, Cumo F, Tiberi M, Sforzini V, Piras G. Cost-benefit analysis for energy management in public buildings: Four Italian case studies. *Energies (Basel)* 2016;9:1–17. <https://doi.org/10.3390/en9070522>.
- [15] Putra HP, Hariana, Nuryadi AP, Nainggolan LMT, Nielsen C. Indonesian Coal Combustion Characteristics Using TG-DSC Analysis. *Proceedings of the International Conference on Innovation in Science and Technology (ICIST 2020)* 2021;208:114–7. <https://doi.org/10.2991/AER.K.211129.026>.
- [16] Siva Reddy V, Kaushik SC, Tyagi SK. Exergetic analysis and evaluation of coal-fired supercritical thermal power plant and natural gas-fired combined cycle power plant. *Clean Technologies and Environmental Policy* 2013 16:3 2013;16:489–99. <https://doi.org/10.1007/S10098-013-0647-X>.
- [17] Zhakupov D, Kulmukanova L, Sarbassov Y, Shah D. Flue gas analysis for biomass and coal co-firing in fluidized bed: process simulation and validation. *Int J Coal Sci Technol* 2022;9:1–11. <https://doi.org/10.1007/S40789-022-00531-Y/FIGURES/3>.
- [18] Al-Lagtah NMA, Al-Habsi S, Onaizi SA. Optimization and performance improvement of Lekhwair natural gas sweetening plant using Aspen HYSYS. *J Nat Gas Sci Eng* 2015;26:367–81. <https://doi.org/10.1016/J.JNGSE.2015.06.030>.
- [19] Oh SY, Kim JK. Operational optimization for part-load performance of amine-based post-combustion CO₂ capture processes. *Energy* 2018;146:57–66. <https://doi.org/10.1016/J.ENERGY.2017.06.179>.
- [20] Braakhuis L, Knuutila HK. Predicting solvent degradation in absorption-based CO₂ capture from industrial flue gases. *Chem Eng Sci* 2023;279:118940. <https://doi.org/10.1016/J.CES.2023.118940>.
- [21] Wilson DGTK. The Design of High-Efficiency Turbomachinery and Gas Turbines n.d. <https://mitpress.mit.edu/9780262526685/the-design-of-high-efficiency-turbomachinery-and-gas-turbines/> (accessed August 31, 2023).
- [22] Indonesia Coal Price | JWC Indonesia Energi n.d. <https://www.jwcindonesia.com/price-index> (accessed May 29, 2024).
- [23] Monoethanolamine price index - businessanalytiq n.d. <https://businessanalytiq.com/procurementanalytics/index/monoethanolamine-price-index/> (accessed May 29, 2024).
- [24] Water Price | PUB, Singapore's National Water Agency n.d. <https://www.pub.gov.sg/Public/WaterLoop/Water-Price> (accessed May 29, 2024).
- [25] Schmitt T, Leptinsky S, Turner M, Zoelle A, White CW, Hughes S, et al. Cost and Performance Baseline for Fossil Energy Plants Volume 1: Bituminous Coal and Natural Gas to Electricity 2022. <https://doi.org/10.2172/1893822>.

- [26] Cost Indices – Towering Skills n.d. <https://toweringskills.com/financial-analysis/cost-indices/> (accessed May 29, 2024).
- [27] Richard Turton, Richard C. Bailie, Wallace B. Whiting JAS. Analysis, Synthesis and Design of Chemical Processes Third Edition. vol. 53. 2013. <https://doi.org/10.1017/CBO9781107415324.004>.
- [28] Seider WD, Lewin DR, Seader JD, Widagdo S, Gani R, Ng KM. Product and process design principles. Fourth edi. New York: Wiley; 2017.
- [29] Couper JR, Penney WRoy, Fair JR, Walas SM. Chemical process equipment: selection and design. Rev. 2nd e. London: Elsevier Butterworth-Heinemann; 2009. <https://doi.org/https://doi.org/10.1016/C2011-0-08248-0>.
- [30] Schoen HM. Plant Design and Economics for Chemical Engineers. Max S. Peters. McGraw-Hill, New, York, 1958. xi+ 511 pp. Illus. \$11. Science (1979) 1958;128:1000–1000. <https://doi.org/10.1126/SCIENCE.128.3330.1000.B>.
- [31] Find minimum of function using pattern search - MATLAB patternsearch n.d. <https://www.mathworks.com/help/gads/patternsearch.html> (accessed April 28, 2023).
- [32] GenOpt n.d. <https://simulationresearch.lbl.gov/GO/> (accessed August 10, 2021).
- [33] MATLAB - MathWorks - MATLAB & Simulink n.d. <https://www.mathworks.com/products/matlab.html> (accessed January 29, 2021).
- [34] Narayanan M, de Lima AF, de Azevedo Dantas AFO, Commerell W. Development of a coupled TRNSYS-MATLAB simulation framework for model predictive control of integrated electrical and thermal residential renewable energy system. Energies (Basel) 2020;13:1–29. <https://doi.org/10.3390/en13215761>.
- [35] Suresh MVJJ, Reddy KS, Kolar AK. 3-E analysis of advanced power plants based on high ash coal. Int J Energy Res 2010;34:716–35. <https://doi.org/10.1002/ER.1593>.
- [36] Oi LE. Aspen HYSYS Simulation of CO₂ Removal by Amine Absorption from a Gas Based Power Plant. SIMS2007 Conference 2007. <https://ep.liu.se/ecp/027/008/ecp072708.pdf> (accessed January 14, 2025).
- [37] Lawal A, Wang M, Stephenson P, Obi O. Demonstrating full-scale post-combustion CO₂ capture for coal-fired power plants through dynamic modelling and simulation. Fuel 2012;101:115–28. <https://doi.org/10.1016/J.FUEL.2010.10.056>.
- [38] Shirdel S, Valand S, Fazli F, Winther-Sørensen B, Aromada SA, Karunarathne S, et al. Sensitivity Analysis and Cost Estimation of a CO₂ Capture Plant in Aspen HYSYS. ChemEngineering 2022, Vol 6, Page 28 2022;6:28. <https://doi.org/10.3390/CHEMENGINEERING6020028>.
- [39] Supekar SD, Skerlos SJ. Reassessing the Efficiency Penalty from Carbon Capture in Coal-Fired Power Plants. Environ Sci Technol 2015;49:12576–84. https://doi.org/10.1021/ACS.EST.5B03052/SUPPL_FILE/ES5B03052_SI_001.PDF.
- [40] Karimi M, Hillestad M, Svendsen HF. Capital costs and energy considerations of different alternative stripper configurations for post combustion CO₂ capture. Chemical Engineering Research and Design 2011;89:1229–36. <https://doi.org/10.1016/J.CHERD.2011.03.005>.