CO_2 Capture Technologies and Shortcut Cost Correlations for Different Inlet CO_2 Concentrations and Flow Rates. Part 1: Chemical Absorption

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Abstract

Carbon capture is a fast-growing sector with increasing commercial interest, yet significant uncertainties remain regarding its real performances. While many publications report CO_2 capture costs, which is one of the key performance indicators (KPIs) for capture technologies, there is a lack of harmonized cost estimation methods, making it difficult to compare technologies on a consistent basis. This has led to considerable discrepancies and inconsistencies across the literature. Moreover, capture costs strongly depend on process parameters such as CO_2 concentration and flue gas flow rates, which results in capture costs across the literature often being incomparable. To address these issues, this paper presents a framework for fairly evaluating the capture cost of chemical absorption technology via shortcut cost correlations. Amine scrubbing is selected as a benchmark technology, with comprehensive validations to precisely depict capture costs across the ranges of CO_2 concentrations (5–50 mol%) and feed flow rates (equivalent to a capture scale of 31–1250 kt/y). The proposed correlations can also handle different economic assumptions including utility costs, and estimation methodologies to accurately reflect final capital expenditures (CapEx) and operating expenditures (OpEx) under various industrial scenarios and operating conditions. The results of such easy-to-use correlations can serve as an important KPI in decision-making processes where CO_2 capture implementation costs need to be rapidly assessed.

Keywords: Carbon Capture Technology, Chemical absorption (MEA), Techno-Economic Assessment (TEA), Shortcut Cost Correlations

1. Introduction

Carbon capture (CC) is widely recognized as a critical step in fulfilling climate targets to reduce carbon emissions. In spite of this broad acceptance and its technological advancement, CC has not yet been implemented to the extent that would have been expected a decade ago, indicating rapid actions are needed (IEA, 2019). Nevertheless, the development of carbon capture technologies is a fast-growing field with increasing attention (GCCSI, 2023), and many publications report the cost of carbon capture, which is one of the key performance indicators (KPIs) of capture processes (IEAGHG, 2019; IEA, 2020; Lyons et al., 2021; NETL, 2022).

CCS (Carbon capture and storage) involves the capture, compression, transportation, and geological storage of CO_2. Alternatively, CO_2 can also be re-used, leading to CCU (Carbon capture and use). There are a number of approaches for CO_2 capture, including post-combustion, pre-combustion, oxy-fuel combustion, and Direct Air Capture (DAC) systems (Osman et al., 2021). The separation and capture of CO_2 from flue gas generated by burning fossil fuels in the air is referred to as post-combustion capture. In the case of pre-combustion, CO_2 is captured prior to fuel combustion,

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where the fuel is gasified with oxygen or air to produce syngas containing CO_2 and H_2 at high pressure. Oxy-fuel combustion involves burning fuel in a pure O_2 stream to create flue gas that is mostly composed of CO_2 and H_2O where CO_2 is then separated via condensation. Under each combustion method of carbon capture, there are many technology options available such as amine solvents, membranes, pressure/ temperature swing adsorption (PSA and TSA), DAC as well as hybrid technologies where multiple capture options are combined (e.g. a membrane technology with an absorption system). An overview is presented in Figure 1.

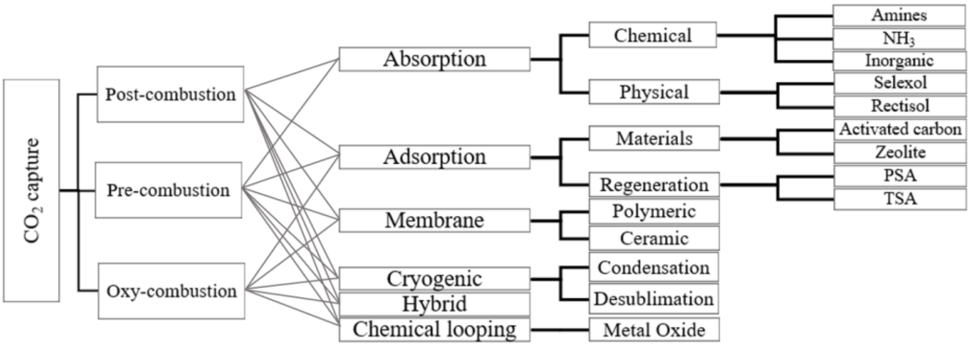

Figure 1. Carbon capture option classification

However, not all technologies are equally appealing and feasible due to factors such as applications, technical/economic limitations, and CO_2-containing gas conditions. In terms of compositions, CO_2 concentration can typically vary between 0.04% (e.g. in the atmosphere) and more than 76% (e.g. oxy-combustion flue gas), and in some specific cases such as ethylene oxide production and anaerobic fermentation, it can be close to 100%. Also, the pressure of flue gas can range from 1 bar to 80 bar depending on emission sources (Zhang et al., 2020).

There are many organizations, including DOE/NETL, IEA, GCCSI, EPRI, etc., and numerous works of literature evaluating the feasibility of capture technologies. However, most of the studies focused on flue gases with a constant CO_2 composition at a fixed capture scale (i.e. the amount of CO_2 to be captured per year), while these parameters heavily impact the economics of the CO_2 capture process (Hasan et al., 2012; Husebye et al., 2012; Zhang et al., 2020). Leeson et al. (2017) reviewed 164 articles related to industrial CC and mean avoidance costs (capture cost including additional CO_2 generated due to the implementation of CO_2 capture units) and the minimum and maximum ranges of avoidance costs from the literature on different industries are summarized in Table 1.

Table 1. Mean avoidance cost ($US\$_{2013}$) of CC technologies across the industries (Based on Leeson et al., 2017)

	Post Combustion (MEA)		**Calcium Looping**		**Mineral Carbonation**		**Oxy-Combustion**	
Industry	Cost	Min/ Max	Cost	Min/ Max	Cost	Min/ Max	Cost	Min/ Max
Iron & Steel	74	20/ 120	N/A	0	62.5	10/ 118	58	N/A
Refineries	78	40/ 180	88	62/ 130	N/A	0	84	62/ 130
Pulp & Paper	58	56/ 59	N/A	0	N/A	0	N/A	0
Cement	107	62/ 164	37.78	20/ 78	N/A	0	59.46	N/A

The biggest cost difference between the minimum and maximum avoidance costs observed in the literature was coming from amine technology for the cement industry (Leeson et al., 2017). It is important to emphasize that each avoidance cost presented by Leeson et al. (2017) is based on a specific fixed capture scale and CO_2 inlet concentration. In general, such literature-based costs can assist as valuable data to identify the magnitude of expected costs. However, when both concentration and scale need to be varied simultaneously, accurately identifying expected capture (or avoidance) cost from the vast amount of publications and making a technology selection would be difficult as there are many layers of information, assumptions, and calculations involved with the capture costs.

Furthermore, another cause of inconsistencies in the capture costs is due to the differences in cost estimation methods and underlying assumptions used in techno-economic assessments (TEAs). As Rubin (2012) and many authors (Van der Spek et al., 2019; Roussanaly et al., 2021; Aromada, 2022) pointed out, there are no harmonized cost estimation methods in the literature yet and many studies fail to pay close attention to the methods and assumptions used in TEAs, resulting in many capture cost results being incomparable and case-specific. Also, many authors often compare the final capture costs with literature-based capture costs without providing detailed breakdowns of the elements used to calculate the capture cost where there are many factors influencing the capture costs. The scarcity of published data from industrial CO_2 capture units also leads to many papers referring to each other, with real-life validation missing. This may however lead to a potential bias in cost estimates for CO_2 capture units in the scientific literature that will be discussed further in this paper.

Therefore, this paper presents a framework to fairly estimate capture costs and the impact of CO_2 concentration and capture scale via shortcut correlations. This framework is intended to assist with technology selection processes and in the present paper, the focus is made on an amine-based carbon capture technology for which technical parameters (e.g. specific-energy consumptions) and economic parameters including total equipment cost (TEC) are evaluated across a wide range of feed gas CO_2 concentration and flow rate. The novel features of this study include the following: (1) cost correlations describing TEC based on detailed cost validations are presented where different estimation methods and location factors can be applied to obtain case-specific final capital expenditures (CapEx); (2) variable operating costs associated with energy consumptions are decoupled from the total operating costs (OpEx) to allow flexibility with the fluctuating utility costs to reflect more accurate capture cost under various scenarios. It is worth mentioning that the outcomes of the correlations are still estimations and there will always be a question about the uncertainty. However, these correlations aim to provide a framework where assumptions and estimation methods are transparently disclosed and therefore, implementation of such correlations is one step forward in establishing a clear and easy-to-implement estimation technique in the public domain. The structure of this paper first introduces cost estimation methodologies for carbon capture technologies in the literature in section 2 while the process and economic parameters used in this study are presented in section 3. Then methodologies for correlation development for the amine capture technology are presented in section 4, followed by results and case studies using the correlations developed in this work.

2. Current status and challenges on CC costing methodologies

When calculating capture costs or comparing results from different literature, it is important to understand project scopes, terminology, estimation methods, assumptions, and the items included and/or excluded from the studies as different aforementioned factors can have a significant impact on the overall outcomes. In this section, potential sources of errors and challenges in the costing methods are discussed.

2.1 System boundary and scopes

Before performing any TEAs or cost estimation studies, it is vital to clearly state the scope and boundaries. Depending on the boundaries of a study, the definition of capture costs may vary; there are multiple steps where CO_2-containing gas can be pre-treated to remove any impurities and then fed through CO_2 capture processes, which consists of the separation of CO_2 from other gases. The captured CO_2 stream can then be post-treated, for instance by removing additional impurities, by compressing it, or by liquefying it for transport. Finally, CO_2 can be injected underground and stored, or it can be re-used.

There are different classifications for plant-level cost estimates defined by the Association for the Advancement of Cost Engineering International (AACE), ranging from class 1 (final detailed study) to 5 (conceptual design) and presented in Table 2.

Table 2. Estimate class defined by AACE (AACE, 2011)

Estimate class	Maturity level of project definition	Purpose of study	Methodology	Expected accuracy: typical +/- range [a]	Preparation effort [b]
Class 5	0% - 2%	Feasibility/ screening	Factors/ models (or judgement)	4 to 20	1
Class 4	1% - 15%	Feasibility/ Concept study	Primarily based on factors/ models	3 to 12	2 to 4
Class 3	10% - 40%	Budget authorisation/ control	Mixed but primarily based on factors/ models	2 to 6	3 to 10
Class 2	30% - 75%	Control/ bid	Primarily deterministic	1 to 3	5 to 20
Class 1	65% - 100%	Estimation check/ bid	Deterministic	1	10 to 100

[a] The index value of '1' indicates +10/-5% and '10' represents +100/-50%
[b] The index value of '1' represents 0.005% of project cost and '100' represents 0.5%

The classes in Table 2 indicate the level of project maturity and the levels of effort required as a project advances from a simple study to the final stages of design and construction. Rubin (2012) stated that the most CC costs available in the literature can be classified as class 3 – 5 while the results of class 1 studies from FEED (Front-End Engineering Design) are often not available in the public domain. Given this information, some levels of uncertainties can be expected from various reports/literature available publicly. In order to establish consistent and standardized estimation methods, several international organizations have developed detailed guidelines (assumptions) and terminologies for calculating plant-level CapEx and OpEx individually (EPRI, 1993; IEAGHG, 2009; DOE/NETL, 2011; CAESAR, 2011). However, many researchers emphasized that the results from these organizations and publications are rather confusing than clarifying, and not comparable without significant efforts (Rubin, 2012; Rubin et al., 2013; Skagestad et al., 2014; Van der Spek et al., 2019; Aromada, 2022).

2.2 Cost estimation methods: CapEx

The process of designing and sizing equipment is the foundation of CapEx [M€] estimation. CapEx and OpEx are significantly impacted by the cost of major equipment needed for every industrial plant. There are three methods of estimating equipment costs and these are listed in the order of accuracy levels: (1) getting a quote or the equipment's cost directly from the original equipment manufacturer (OEM) or vendor; (2) using the cost of similar equipment (of same kind, and update for construction material, size, and capacity) that was purchased in the previous or recent years; (3) estimating via costing equations/graphs in engineering textbooks and publications. Depending on the level of detail of the available data, it can be further classified as the bottom-up or top-down approaches. For the bottom-up approach, sizing of each piece of equipment in a capture process is obtained via process modelling, and corresponding costing data is applied to estimate each item in equipment costs. The top-down approach, in contrast, indicates the use of historical data of an entire capture process to quickly assess the feasibility of a project.

As mentioned before, there are many organizations and institutions that developed total CapEx estimation methods to calculate initial cost estimations based on the TEC which is the sum of all process equipment costs within a defined process boundary while the other costs associated with CapEx are estimated as factors of the TEC (Ali et al., 2019; IEAGHG, 2009; DOE/NETL, 2011; Rubin et al., 2013; Aromada, 2022). However, since there are no harmonized guidelines for carbon capture cost estimation methods, each guideline tends to employ different terminologies and can include (or exclude) some cost elements in different ways. Table 3 presents some examples of many existing methods in literature and common nomenclatures used in different organizations (Peters et al., 2018; CAESAR, 2011; DOE/NETL, 2011).

Table 3. Capital cost estimation methods and nomenclatures (adapted from Rubin et al., 2013)

Peters et al., (2003)	% of PEC	Sum	CAESAR (2011)	% of TEC	Sum	DOE/NETL (2021)	% of TEC	Sum
Direct Cost (Process type: Fluids)			**Direct Cost**					
ISBL								
- Purchased Equipment Cost (PEC)	100		- Total Equipment Cost (TEC)	100		**Total Equipment Cost (TEC)**	100	
- Installation	47		- Erection, steel structure and painting	49		Supporting Facilities (Turton et al., 2018)	71.4% TEC	
- Instrumentation/ Control	18		- Instrumentation/ Control	9		Total Labor Cost (Turton et al., 2018)	37% (TEC + 71.4 TEC)	
- Piping	66		- Piping	20		**Bare Erected Cost (BEC)**	Sum of above	234.8
- Electrical	11		- Electrical Equipment and materials	12				
OSBL			- Civil works	11				
- Building and building services	18		**Total Direct Plant Cost (TDPC)**	Sum of above	201			
- Yard improvements	10							
- Services facilities	70							
- Land	6							
Total Direct cost (ISBL + OSBL)	Sum of above	346						
Indirect Cost			**Indirect Cost**					
Engineering	33		- Engineering and Supervision	13.89		- Engineering and Construction management	15-20% of BEC = 41% of TEC	
Construction Expenses	41		- Buildings (with services)	8.33		**Engineering/Procurement/ Construction (EPC)**	Sum of above	275.8
Contractor's Fee	21		- Service facilities	4.20				
Project/ Process contingency	42		- Yard improvements	2.78		Process Contingency	20% of EPC = 55.2% TEC	
Total Indirect cost	Sum of above	137	**Total Indirect cost**	Sum of above	29.20	Project Contingency	15-30% of (BEC + EPC + 25% of BEC) = 128.1% TEC	
Fixed Capital Investment (FCI)	Sum of above	483	**Engineering/Procurement/ Construction (EPC)**	TDPC + Indirect cost	230.20	**Total Plant Cost (TPC)**	Sum of above	459.1
Working Investment	86		Owner's cost/ Contingency	34.5		Owner's cost	Up to 20% TPC = 91.8% TEC	
						Start-up/ Inventory		
Total CAPEX	Sum of above	**569**	**Total Plant Cost (TPC)**	Sum of above	**264.70**	**Total CAPEX**	Sum of above	**550.9**

*ISBL (Inside battery limit), OSBL (Outside battery limit): includes building and services, yard improvements, service facilities, land

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Rubin et al. (2013) noted that although the structural frameworks of cost estimation methods are similar, the specific items considered within each category can differ, leading to discrepancies of over 100% in the total capital expenditure (CapEx) estimates for similar total equipment costs (TEC). The percentage factors applied to each category vary between methods, as illustrated in Table 3, which can result in significant variations in the final estimates. For instance, Rubin (2012) highlighted that incorrect process and project contingencies could lead to a CapEx underestimation of 50% or more. Additionally, discrepancies can arise from the treatment of owner's costs in various methods, with differing definitions and inclusions. Terminology such as Total Plant Cost (TPC) can be misleading to the general public, as it suggests that all associated costs are included, while the definition of TPC varies across methodologies. Consequently, a considerable amount of time and effort is required to achieve fair comparisons of capture costs from the literature. It is also important to note that the methods discussed have not been updated for decades, and their applicability requires urgent reassessment (Van der Spek et al., 2019).

Apart from the structural differences in the CapEx estimation methods, guidelines to quantify some items such as total equipment costs, supporting facilities, and labour costs are often not given since these methods are based on costs prepared by contractors and technology providers. Full transparent details/ guidelines of such methods (e.g. percentages accounted for supporting facilities and labour costs in DOE/NETL method are not described) are often not disclosed to the public. As a large company, it is a common practice to employ highly experienced specialists to collect cost data and work with the vendors given project budgets. Also, in reality, actual equipment costs may vary significantly depending on the required quantities and the timelines of the projects (Sinnott and Towler, 2008). These indicate that if high-quality cost data is not available when conducting TEAs, engineers often have no other option but to select less accurate alternatives such as guidelines/ estimation equations from textbooks to quantify items to complete cost studies.

As a result, it was commonly observed that the capital cost estimation methods are intermixed with each other and some elements especially calculating equipment costs are estimated via heuristics and cost equations/ graphs published in textbooks (Richardson, 2002; Sinnott and Towler, 2008; Turton et al., 2018) depending on the availability of data (Nazir et al., 2018; Nwaoha et al., 2018; Wang, 2020). The choices made in equipment costs and total CapEx estimation methods can heavily influence the final outcome as presented in Table 3 and therefore, it is emphasized that transparency is very important to provide coherent cost estimates that are comparable.

Besides various methods to calculate equipment costs and CapEx mentioned above, there are additional sources of errors and uncertainties in TEA studies: (1) Cost index; (2) location factors; (3) currencies. In the case of cost indices, many studies implement the US Chemical Engineering Plant Cost Index, also known as the CEPCI (Chemical Engineering Magazine, 2023). However, as the name shows, it's applicable to the United States and may not accurately present other regions or countries. Also, other indices such as the Dutch and German indices show a clear deviation from the US CEPCI, and therefore, the selection of an index can result in different estimations (Van der Spek et al., 2019). For location factors, organizations such as IEAGHG, (2018) and GCCSI, (2011) as well as authors (Garrett, 1989; Richardson, 2002; Brown, 2019) published correction factors to account for cost differences in locations and

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labour costs as well as productivity. It is common to observe significant differences in equipment, material, and labour costs across countries and it is important to consider it when conducting TEA studies in different locations. Lastly, simply converting the currency can cause bias in the results. Overall, the principles of cost estimation methods are straightforward; however, the aforementioned factors make the results obtained from various CapEx methods incomparable and challenging to understand without detailed analysis.

2.3 Cost estimation methods: OpEx

The elements of OpEx [M€/y] commonly considered in various estimation methods can be categorized into fixed and variable O&M (operating and maintenance) costs. A general breakdown of OpEx can be found in Table 4 (IEAGHG, 2013). The fixed O&M costs consist of mainly labour and maintenance costs which are independent of plant utilization while variable O&M costs are directly proportional to the utility consumptions and pollution controls. Rubin et al. (2013) observed that some methods treat maintenance materials as variable cost items but in general for OpEx, the methods are similar across the existing organizations. However, similar to what was discussed in CapEx methods in section 2.2, some OpEx methods do not provide guidelines to estimate all cost items as a percentage of reference costs in general. Indeed, some costs depend on case-specific parameters such as fuel price, utility costs, labour wage, etc. For the missing items in OpEx estimation methods, heuristics and assumptions found in engineering textbooks such as Richardson, (2002), Sinnott and Towler, (2019), and Turton et al. (2018) are commonly used in the literature. After estimating the CapEx and OpEx of a capture process, the capture cost can be calculated. An overview and calculation methods on the capture cost are presented next.

Table 4. General OpEx elements for carbon capture processes (IEAGHG, 2013)

Cost items to be quantified	Sum of preceding items
Operating labour	
Maintenance labour	
Administrative and support labour	
Maintenance materials	
Tax/ Insurances	
	Fixed O&M costs
Fuel/ Electricity	
Other consumables:	
E.g. Catalyst, chemicals, auxiliary fuels, water	
Waste disposal (excluding CO_2)	
By-product sales/ emission tax	
	Variable O&M Costs

2.4 CO_2 capture cost

When it comes to the capture costs, it is important to analyze and compare the methods and input assumptions considered in the TEAs rather than directly comparing the capture cost results. However, in the literature, a common practice is to compare CO_2 capture costs with other publications which is not enough to identify and justify the discrepancies in the cost results. Also, there are three ways to calculate CO_2 capture costs (Roussanaly, 2019). The first one is the ‘exhaustive’ method in which the capture (or avoidance) cost is based on the process cost and emission intensity of the key material(s) (e.g. the main product) of an industrial plant with and without a carbon capture process. This approach requires more details and time but would be always valid regardless of the scenarios. An example is the case of hydrogen production with CC where carbon capture and production of H_2 take place in the same unit and therefore, a detailed ‘exhaustive’ method should be used. The second method is called the ‘net present value’ method which is based on all future cash flows of a CC investment over the entire project life discounted to the present. In contrast to the ‘exhaustive’ method, this approach is independent of the cost of a plant without carbon capture. However, one needs to check that the performance of the industrial plant is not affected by the CC implementation and the cost associated with the CC unit must be independent of the plant. The third option is the ‘annualization’ method which is presented in Equations 2 and 3 where the annualized CC investment is divided by the annual amount of captured (or avoided) CO_2. In this paper, carbon capture technologies are assumed to be independent of the connected industrial processes. This means that the production of the industrial plant of interest will not be affected regardless of the presence of a carbon capture unit. Also, OpEx and CO_2 emissions from carbon capture technologies are assumed to be constant throughout the operation. Therefore, under these assumptions, the use of the ‘annualization’ method is justified. Nevertheless, depending on the functionality and interactions between the existing plant and the CO_2 capture plant, different methods should be implemented. An overview of elements used to estimate the capture cost in the annualization method is presented in Figure 2.

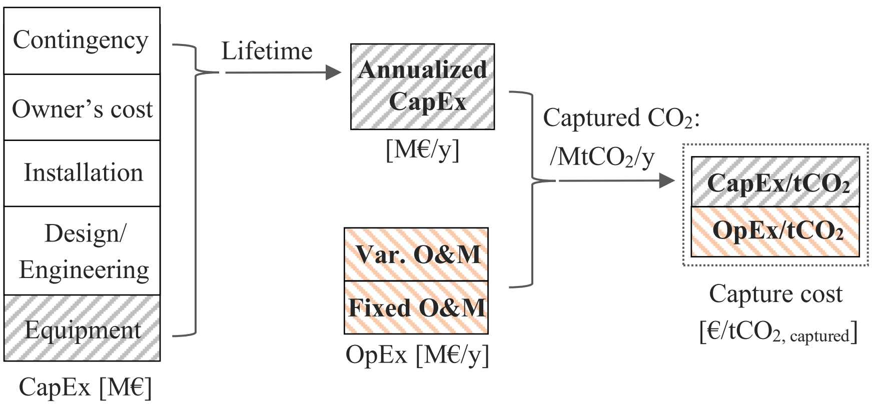

Figure 2. General scheme for carbon capture cost calculation

After obtaining the sum of equipment costs, one of the CapEx methods presented in section 2.2 can be implemented to account for items such as installation, engineering, contingencies, etc., and the sum of all the items is then referred to as the total CapEx [M€]. Over the lifetime of the capture technology (usually 30 years but may vary depending on the technology), the CapEx is annualized [M€/y] using the following equation:

$$Annualized\ CapEx = CapEx\left(\frac{i(1+i)^n}{(1+i)^n - 1}\right) \quad \text{Eq. (1)}$$

Where i is the discount rate and n is the lifetime of a capture process. OpEx [M€/y] is then added to the annualized CapEx and the sum is divided by the total amount of CO_2 captured [Mt of CO_2/y] over an operating year.

$$CO_2\ capture\ cost = \frac{Annualized\ CapEx + OpEx}{CO_{2,\ captured}} \quad \text{Eq. (2)}$$

When the process emissions from the capture process itself are deducted from the total amount of CO_2 captured in Equation 2, CO_2 avoidance cost can be obtained.

$$CO_2\ avoidance\ cost = \frac{Annualized\ CapEx + OpEx}{CO_{2,\ captured} - capture\ process\ CO_2\ emission} \quad \text{Eq. (3)}$$

Both CO_2 capture cost and CO_2 avoidance cost (or CO_2 avoided cost) are common metrics used in CC literature. The capture cost presents investment and operating costs involving the capture unit while the avoidance cost provides the performance of the capture unit as a mitigation option (Roussanaly et al., 2021). In this paper, only CO_2 capture costs are presented to simplify the considered case studies. Lastly, another important aspect to check when calculating capture cost is utility costs and their generation sources used in a TEA. For example, Roussanaly et al. (2017) studied the costs of steam generated via several sources and these differences in the steam costs can significantly influence the final capture cost by up to 37%.

As can be seen in the above sections, there are many aspects (CO_2 concentration, flow rates of the feed gas, economic assumptions, estimation methods, etc.) that need to be carefully considered to calculate the cost of carbon capture and it can be a time-consuming task. To greatly reduce the cumbersomeness of detailed TEA studies, there are a few published cost correlations that directly relate capture costs to some process input parameters.

2.5 Pitfalls in cost correlations in the literatures

In order to address the impacts of process conditions on capture costs, there have been some efforts to study the influences of concentrations and/or flow rates on the economics of capture processes via correlations (Hasan et al., 2012; Pieri and Angelis-Dimakis, 2021; Zhang et al., 2020). In the work of Pieri and Angelis-Dimakis, (2021), data points used in correlations were collected across the literature. When OpEx costs were not available, these values were estimated by assuming a percentage of reported capital costs. It is important that these data points were standardized to the same costing year of 2018, the same currency (US $), and the system boundaries were adjusted to consider only capture plants if applicable. These harmonized data were then used to develop correlations in the following format:

$$\mathrm{y} = \mathrm{a} \cdot \dot{m}^{b} \quad \text{Eq. (4)}$$

Where a and b are fitting parameters and $\dot{m}$ is the captured CO_2 [Mt/y] and y can be either total capital requirements or operating costs [M$/y]. Despite the fact that these literature-based data were uniformized to develop the correlations, establishing a fair comparison using such correlations would be complicated as the effects of flue gas concentration on economic parameters are neglected in the correlation. In the work of Zhang et al. (2020), chemical, physical, and hybrid absorption systems are simulated to obtain data points for regressions. The authors follow a correlation format as a function of CO_2 concentration (x_{co_2}) which was derived from Hasan et al. (2012), without including the influence of flow rates:

$$\text{Total Annual Cost (TAC)} = \alpha + \left(\beta \cdot x_{co_2}\right)^{\gamma} \quad \text{Eq. (5)}$$

Where TAC [M$/y] is the sum of annualized CapEx and OpEx, and α, β and γ are the fitting parameters. Lastly, the correlation of Hasan et al. (2012) utilizes both concentrations and flow rates (*n* and *m* are fitting parameters) to predict annualized CapEx and operating cost (OpEx) independently:

$$Z = \alpha + (\beta \cdot x_{co_2}{}^{n} + \gamma) \cdot F^{m} \quad \text{Eq. (6)}$$

However, the correlations of Hasan et al. (2012) lack the freedom to implement other cost estimation methodologies presented in section 2.1 since the final outcome of the correlations are annualized CapEx and OpEx. Also, users are not able to introduce alternative utility costs which heavily influence the capture cost (Roussanaly et al., 2017). These aforementioned correlations are useful to obtain a good order of magnitude for capture costs but when there are new conditions and input data are available, there is a gap in the flexibility of the correlations which fails to reflect the capture cost accurately. In this work, the correlation format of Hasan et al. (2012) is adopted and further improved by allowing flexible applications of the correlation outcomes. The amine capture process is modelled in Aspen Plus with a literature-based benchmark for technical and economic validations. Then the model is used to generate data points to develop correlations at various CO_2 inlet concentrations and flue gas flow rates. More details about the correlation development are presented in section 4. The process and economic parameters used in the Aspen Plus models are described next.

3. Process and Economic Parameters

The boundary of cost estimations includes CO_2 capture and compression systems. The cost associated with pre-treatments, transportation and sequestration/utilizations are not included in the system boundary as presented in Figure 3. Although many sources of CO_2-containing gases often carry impurities such as NO_x and SO_x, it is assumed that the inlet feed gas is free of impurities. Additionally, the temperature of the flue gas is set at 170°C to prevent acid gas condensation. In general, flue gases contain 5.5-15% of H_2O (Hasan et al., 2012), and in this work, 11% of H_2O is chosen as a default value (all gas composition are given in volume percentages). Typical content for O_2 is 2-8% and 6.4% is used in this study, and the balance of the feed gas is assumed to be N_2. Oxygen content may impact the stability of capture solvents, but this effect has been excluded from the scope of the present study and can be considered in future works. The pressure of the feed gas is constant at 1 bar as post-combustion flue gases are available at atmospheric pressure. In Aspen Plus, a feed gas cooler was implemented to reduce the temperature of the stream to 40°C before entering the absorber. A capture rate of 90% was achieved at various CO_2 concentrations and feed gas flow rates and the minimum purity of the CO_2 product stream was set at ≥ 95%. Due to the case-specific nature of waste heat data, heat integration is out of the scope of this paper. However, one of the results of the present work is to address energy requirements, and such outcomes could be further used for heat integration studies. The captured CO_2 is finally compressed to 150 bar for transportation and this occurs via a multi-stage compression system with intercoolers presented in section 3.2. These choices are arbitrary and may impact the results.

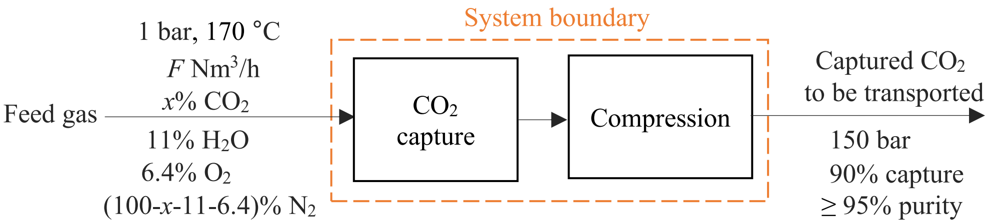

Figure 3. Assumed system boundary for selected capture technologies (pre-treatment, transportation, sequestration/re-use are excluded)

In the present paper, technology-specific equipment costing data/ equations from Nwaoha et al. (2018) are used to calculate equipment costs of the capture process considering amine-based CO_2 capture. This approach is based on ballpark estimates provided by vendors and original equipment manufacturers (OEMs). In Aspen Plus, a model similar to the process described in Nwaoha et al. (2018) is developed and technical and cost validations are performed to ensure the validity of both the process and cost models used in TEAs before generating datapoints for correlation development.

3.1 Chemical Absorption (MEA) Process

Monoethanolamine (MEA) process is one of the widely investigated capture technologies where lean MEA solvent (30 wt.% in water) selectively absorbs CO_2 from flue gases and is then sent to the stripper for solvent regeneration (100-140°C at atmospheric pressure). The regenerated lean MEA solvent is recycled back to the absorber to complete the solvent loop. A general process flowsheet of MEA process is presented in Figure 4.

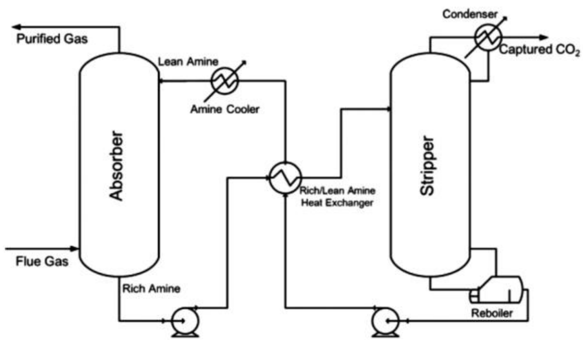

Figure 4. Flowsheet of chemical absorption system (Kazemifar, 2022)

3.1.1 Process and Simulation specifications

In Aspen Plus, the electrolyte-NRTL thermodynamic package with ideal gas law is used and the absorber is designed using equilibrium calculations and then converted to the rate-based calculations to model equilibrium and kinetic controlled reactions in the absorber. A design specification was set in the absorber to achieve 90% capture rate by changing required reboiler duty in the stripper. Aspen Plus MEA model is a complex model as reactive absorption takes place in the rate-based nonconvex column models. The presence of electrolytes in the liquid phase contributes to the convergence difficulty of the model. In addition, the model simulates multiple recycle streams (e.g. lean stream recycles between the absorber and stripper) which increases computing time. To overcome convergence issues, this paper adopted the method presented by Penteado et al. (2016) and Wang et al. (2023) where they implemented a transfer block in Aspen Plus which copies the stream information coming out of the stripper into the inlet lean amine stream to virtually close the recycle streams. A balance block was implemented to calculate the makeup stream flow rates in this work and more details can be found in the above-mentioned literature. Other process parameters are listed in Table 5.

Table 5. MEA process column specifications

Specifications	Values
Absorber	Rate-based calculation
Packing	Sulzer Mellapak 250Y
Number of stages	20
Top section pressure	1.2 bar
Inlet MEA temperature	40°C
Washer	Equilibrium calculation
Packing	Sulzer Mellapak 250Y
Number of stages	2
Stripper	Equilibrium calculation
Packing	Sulzer Mellapak 250Y
Number of stages	8
Reboiler pressure	1.7 bar

In order to design the absorber, design procedure of Agbonghae et al. (2014) was followed where firstly, the lean solvent flow rate required to capture 90% of the CO_2 in the flue gas entering the absorber can be estimated as following:

$$L_{Lean} = \frac{G m_{co_2} \varphi_{co_2}}{100z(\alpha_{Rich} - \alpha_{Lean})} \cdot \left[\frac{M_{MEA}}{44.009}\left(1 + \frac{1 - \omega_{MEA}}{\omega_{MEA}}\right) + z\alpha_{Lean}\right] \quad \text{Eq. (7)}$$

Where G is the mass flow rate of the feed gas (kg/s), m_{co_2} is the mass fraction of CO_2 in the flue gas (%), φ_{co_2} is the capture rate (%), z is the number of molar equivalents/ mole of amine (1 for MEA), α_{Rich} and α_{Lean} are the lean and rich loading (mol CO_2/mol MEA) respectively, M_{MEA} is the molar weight of MEA (kg/kmol) and ω_{MEA} is the MEA weight percentage. The column diameter (D) can be then estimated given the gas flow rate (G) and the superficial velocity of the gas (U_s) where more details can be found in Agbonghae et al. (2014) and Otitoju et al. (2023):

$$D = \sqrt{\frac{4G}{\pi U_s}} \quad \text{Eq. (8)}$$

To identify the cost optimum height of the column, the height of the absorber is varied at fixed column diameter and solvent flow rate (obtained using Eq. 7) to minimize the TAC of the column which includes equipment cost of the absorber and steam cost for regeneration. Then another sensitivity study on the solvent flow rate is performed at fixed column dimensions to further optimize the reboiler duty which dominates energy consumptions in a typical MEA process. More details of the sizing method can be found in Appendix C.

For the lean/rich heat exchanger, design calculation mode was used in Aspen Plus with a minimum approach temperature of 10°C, and the temperature difference between the hot outlet stream and cold inlet stream set to 10°C. The rich solvent pump in the model was set at 75% efficiency, and the feed gas blower's isentropic efficiency was set at 85%.

3.1.2 Economic data and assumptions

The sizing results from the Aspen Plus model were extracted to estimate the total equipment cost and operating costs. As mentioned before, there are various methods to calculate equipment costs. In this study, vendor/ original equipment manufacturer (OEM) estimates (Table. 6) from Nwaoha et al. (2018) are used to obtain equipment costs. The costs are given in US dollars in the literature above, and therefore, the cost data in Table 6 are assumed to be based in the US Gulf. The uncertainty presented in Table 6 provides an overall uncertainty in the TEC of ± 17.4%.

Table 6. Vendor/ OEM equipment cost and their capacity (Nwaoha et al., 2018)

Equipment	Capacity/Size $S_{i_{ref}}$	Cost (US$ 2018) $C_{i_{ref}}$	Uncertainty (%)
Blower	236 m^3/s	650000	±10
Two-phase vertical separator	2268 kg	105000	±20
Packing [a]	1 m^3	2000	±20
Shell and tube heat exchanger	1000 m^2	401705	±30

[a] Sulzer Mellapak 250Y Structured packing

Then, Eq. 1 Towler and Sinnott, (2012) is used to obtain applicable equipment costs at a required capacity and size.

$$\frac{C_i}{C_{i_{ref}}} = \left(\frac{S_i}{S_{i_{ref}}}\right)^n \qquad \text{Eq. (9)}$$

Where S is the capacity/size of the equipment, C is the equipment cost from the vendor or OEMs while n is the cost exponent for size/capacity correction. The cost exponent of equipment usually varies from 0.3 to 0.84 but similarly to Nwaoha et al. (2018), it was taken as 0.6, known as the six-tenth rule. Subscripts i and i_{ref} are the current equipment and reference equipment respectively. For the absorber and desorber, the column weight was used to estimate the equipment cost, applying the specific cost per unit mass shown in Table 6. For instance, the column cost is calculated based on a pricing of $40 per pound of column mass (i.e., $40 per 0.454 kg). An example of the column weight calculation can be found in Appendix B. The pump cost was not provided by the vendor/OEMs in Nwaoha's paper and therefore, a cost equation from Towler and Sinnott (2012) was used. The CEPCI was then used to update the past costs to the reference time (798.7; July 2023).

OpEx can be classified as fixed O&M and variable O&M as shown in Table 4. For variable O&M, utility costs for steam, electricity, and cooling water presented in Nwaoha et al, (2018) are used. As presented in section 2.3, not all details to estimate OpEx items are provided in the OpEx estimation methods and therefore assumptions presented in Turton et al. (2018) are implemented for fixed O&M costs. Table. 7 shows the assumptions and the items included in the fixed O&M [M€/y]. The operating labour cost (OLC) is calculated based on the operator number of 16 (Turton et al., 2018) and the annual salary of 50 000 $\$_{2018}$ for the benchmark study (Nwaoha et al., 2018) and a cost scaling law with an exponent of 0.25 is implemented to account the OLC at various capture scales (Peters et al., 2018). Turton et al. (2018) presented two methods to estimate the fixed capital investment (FCI), where a factor of 1.18 can be applied to the TEC, and it is called the total module cost (C_{TM}), which is used interchangeably with the FCI in the textbook. When a process involves significant expansion of the site, an extra factor of 0.5 is multiplied by the TEC at the base conditions (which are the equipment costs at the reference pressure and carbon steel as a material of construction) and added to the C_{TM} to estimate the FCI. In this paper, the MEA capture process was assumed to involve a significant expansion of the site, and therefore, both 1.18 and 0.5 factors are applied, while for the CO_2 compression loop (section 3.2), a factor of 1.18 was used as described above, assuming a minor expansion of the site.

Table 7. Fixed O&M cost estimation assumptions (Turton et al., 2018)

Fixed O&M costs	
Operating labour cost (OLC)	/
Supervision	20% of OLC
Laboratory charges	15% of OLC
Maintenance and repair	5% of FCI*
Operating supplies	0.9% of FCI
Local taxes/ insurance	2 % of FCI
Plant overhead expenditure	70.8% of OLC + 3.6 % of FCI
Administrative expenditures	17.7% of OLC + 0.9% of FCI

*The fixed capital investment (FCI) is calculated as described by Turton et al. (2018)

3.2 Compression loop

After capturing CO_2 from the flue gases, the CO_2 product stream is compressed to 150 bar for transportation. Three compressor stages with inter-coolers up to 80 bar (DECARBit, 2011) are implemented in this study (Figure 5) with inter-coolers at each stage set at 28 °C. Then, a pump is implemented to reach the final pressure of 150 bar. The pressure details for each stage are presented as follows:

Table 8. Compression stages and discharge pressure with isentropic efficiencies

Compression stages	Discharge pressure [bar]	Efficiency [%]
1	4.35	85
2	18.65	85
3	80	80

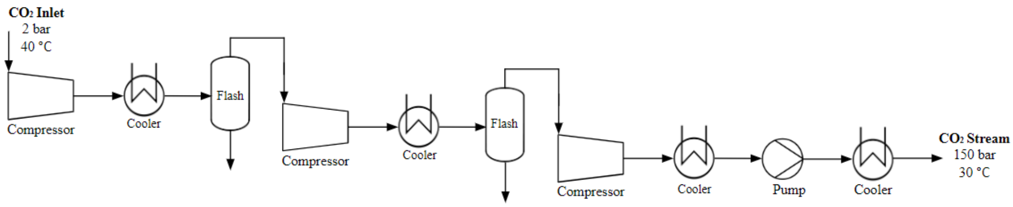

Figure 5. CO_2 product stream compression stages (150 bar, 30 °C)

4. Cost model, data generation and shortcut correlation development

The main goal of this work is to evaluate post-combustion amine-based carbon capture technology with a 90% capture rate at various CO_2 inlet concentrations [mol%] and flow rates [10^3 Nm^3/h], which influences the performances of the capture process and thus capture costs. In the first step, the cost model for amine capture technology is developed using Excel and validated with literature. Then this validated cost model is coupled with the Aspen Plus model where ranges of CO_2 inlet concentrations and flue gas flow rates are evaluated. To do so, a range of concentrations is selected between 5-50 mol% in the inlet flue gas stream, spanning over most typical industrial applications. Capture scales of 31-1250 kt/y were considered in this study to understand the impact of the size of the plant, where for each concentration presented in Table A1, corresponding flue gas flow rates are calculated. The results (such as mass/ energy balances and sizing of equipment) from the Aspen Plus models for each data set are transferred back to the Excel cost model where technical and economic parameters are recorded. These data points are then used in cost correlation development in Matlab where multivariable regressions of correlation parameters are performed. Then the developed correlations can serve as a shortcut model to estimate KPIs (e.g. process energy requirements, final capture costs, etc.) given input parameters such as CO_2 inlet concentration and flue gas flow rate. An overview of the correlation development and correlation framework is presented in Figure 6. The correlation development is an iterative process where more data points can be added to reinforce the correlations.

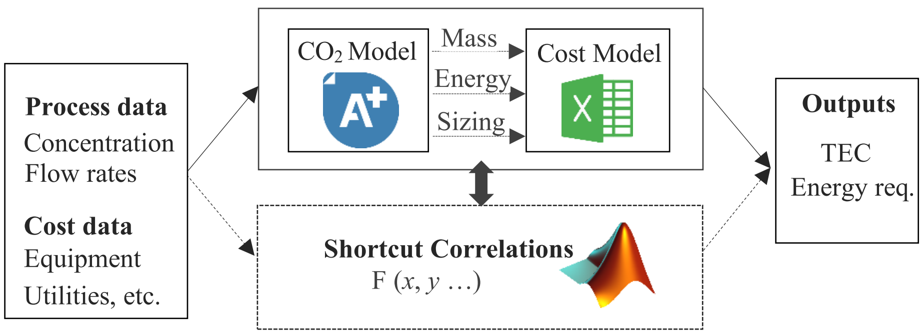

Figure 6. Shortcut correlation model development and framework

The CapEx and OpEx are decoupled into their constituents as shown in Table 9 so that the correlation outputs are now flexible with different assumptions and CapEx/ OpEx methods presented in section 2. Therefore, when the same assumptions, boundaries, and common CapEx/ OpEx methods are applied throughout comparison studies, structural discrepancies found in the estimation methods and inconsistencies in assumptions can be minimized. In addition, one of the reasons why many literature capture cost estimates are not comparable with each other is due to their case-specific utility costs implemented in each TEA. However, in this work, since the total OpEx is decoupled into fixed O&M and specific energy consumptions (heating, electrical, and cooling requirements), case-specific utility costs can be applied to estimate variable O&M costs. This can improve the accuracy of the cost estimation results.

Table 9. CapEx and Total OpEX and related key performance indicators

Parameter	Key performance Indicators	Evaluation method
CapEx	TEC [M€]	Correlation
Total OpEx	Fixed O&M [M€/y]	Turton et al. (2018)
	Variable O&M [M€/y]	Calculated*
	- Specific heat requirement [GJ/t_{CO_2}]	Correlation
	- Specific electrical requirement [kWh/t_{CO_2}]	Correlation
	- Specific cooling requirement [GJ/t_{CO_2}]	Correlation

*Case-specific utility costs can be applied to the specific energy consumptions obtained from the correlations presented in this work.

The correlation for TEC (Eq. 11) is a function of both CO_2 inlet concentration and flue gas flow rates while specific energy requirements are independent of the flue gas flow rates (or the capture scale). Therefore, in this paper, an exponential format is proposed to describe specific energy requirements as a function of CO_2 inlet concentration (Eq. 12).

$$\text{TEC} = \alpha + (\beta \cdot {x_{co_2}}^{n} + \gamma) \cdot F^{m} \qquad \text{Eq. (11)}$$

$$\text{specific energy requirement} = \alpha \cdot e^{n x_{co_2}} + \beta \cdot e^{m x_{co_2}} \qquad \text{Eq. (12)}$$

Note that fixed O&M costs are not described using correlations as a user can select a method to estimate it once TEC and energy requirements are calculated using the correlations presented in this paper. Using the correlations presented above, carbon capture costs at various flue gas conditions (at 1 bar) representing various industrial flue gas streams can be evaluated. The TEC obtained using Eq. 11 at given CO_2 inlet concentration and flue gas flow rate can then be applied to estimate the CapEx following any of the estimation methods presented in section 2.3. In this paper, the DOE/NETL estimation method is selected as a default CapEx costing method since the DOE/NETL method is used by many authors (Element Energy et al., 2014; Nwaoha et al., 2018; Van der Spek et al., 2019; Yun et al., 2021; Aromada, 2022; Langhorst et al., 2022). Also, DOE/NETL is among the few organizations that published detailed cost reports and therefore, validation of economic assumptions and data collection is more convenient

5. Results and discussion

The correlations describing the technical and economic performances of the amine-based carbon capture process are presented in this section. The results of simulation and TEA validations, correlation parameters, and TEA case studies are presented below.

5.1 Simulation and TEC validation

Before generating data for the correlations, the selected carbon capture technology considered in this study was validated in terms of both technical and economic performances. A summary of the MEA capture process implemented in a cement production plant (for the exhaust gas) presented in Nwaoha et al. (2018) is shown in Table 10. Note that all the cost results presented in Nwaoha et al. (2018) were based in 2018 (CEPCI value of 581). In this work, the reference year of 2023 was selected to provide more up-to-date costs of carbon capture, and all the costs presented hereafter (including utility costs in Table 10) were updated using the CEPCI of 2023 July (798.7). An exchange rate of 0.89 was implemented to convert the US Dollar ($) to Euro (€).

Table 10. Process parameters presented in Nwaoha et al. (2018)

Parameters	Values	Parameters	Values
Technology	30% MEA	CEPCI (February 2018)	581
Capture size	0.7 Mt/y	Operating hours [h/y] [a]	8215
Location	Quebec, Canada	Cooling water [$€_{2023}$/GJ]	0.23
CO_2 concentration	11.5 mol%	Low pressure (LP) Steam [$€_{2023}$/GJ]	17.86
Capture rate	90%	Electricity [$€_{2023}$/kWh]	0.056
Discount rate	0.085	MEA price [$€_{2023}$/tonne MEA]	2314

[a] Annual operating hour was re-calculated based on the utility costs and the variable O&M cost.

Table 11 presents the simulation results of the current paper and the technical results as well as typical values found in the work of Nwaoha et al. (2018) for the MEA process.

Table 11. Benchmark literature for MEA and comparison results for validation

	Unit	Simulation Results	Nwaoha et al. (2018)	Typical values
Rich loading	[mol CO_2/mol amine]	0.50	0.50	0.45 – 0.55
Lean loading	[mol CO_2/mol amine]	0.31	0.32	0.15 – 0.36
Reboiler duty	[GJ/t_{CO_2}]	3.76	3.86	3.2 - 5
Solvent flow	[t/day]	48196	48034	NA
Heat exchange area	[m^2]	13617	13433	NA

At similar rich and lean loadings, close specific reboiler duties (2.6% difference) between the simulation results (current study) and the work of Nwaoha et al. (2018) can be observed. The heat transfer coefficient of 732.8 $W/m^2 \cdot °C$ as specified in Nwaoha et al. (2018) was used in the lean/rich heat exchanger and similar heat exchanger areas were observed in both cases as well as the solvent flow rates used in the systems. Besides the technical performances, economic parameters in terms of TEC were then validated and the comparison is presented in Figure 7a. The considered equipment lists for the MEA system are (1) blower, (2) flue gas cooler, (3) rich solvent pump, (4) absorber, (5) stripper, (6) condenser/washer, (7) lean/rich heat exchanger, and (8) lean solvent cooler.

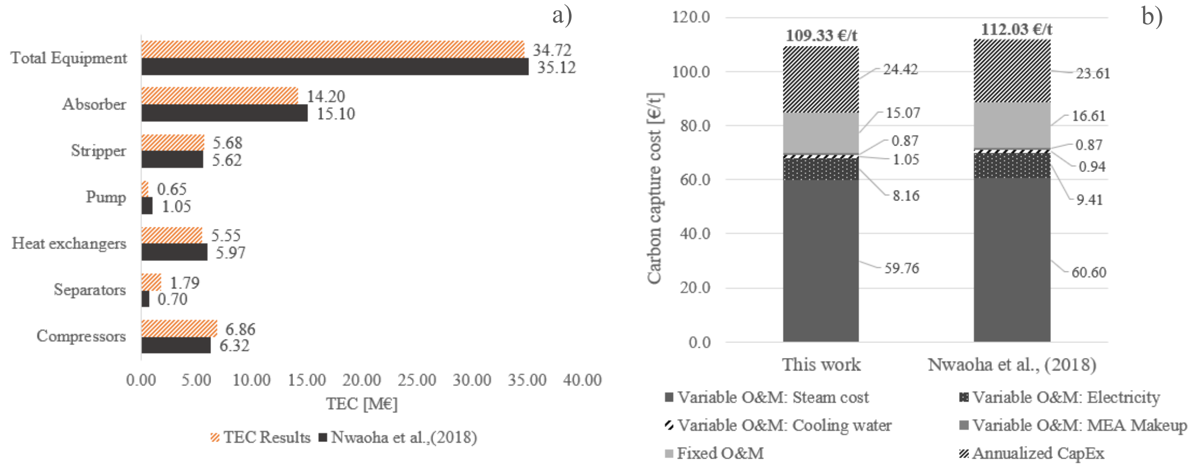

Figure 7. TEC and capture cost comparisons between Nwaoha et al. (2018) and the current work

The largest contributor to the TEC is the absorber which accounts for nearly 40% of the TEC. This is due to the capture system with a huge flue gas flow rate at a high capture rate of 90%. Overall, the differences in TEC were found to be 1.15% which is well within the specified uncertainty of ± 17.4% as well as in agreement with the AACE Class 4 estimate uncertainty of -30% to +50%. A similar trend for equipment items was observed in the studies (Figure 7a). Note that in this paper, within the MEA process system boundary, only one pump was considered in the Aspen Plus model for the rich solvent from the absorber although it appears from the process diagram given in the work of Nwaoha et al. (2018) that multiple pumps are considered in their TEA. Nevertheless, when comparing the results against literature data, it is often not possible to precisely pinpoint differences without comparing the actual simulation models as possible causes for deviations are likely to be different specifications used in the model inputs, absorber/ stripper column mass calculation methods, flowsheet configurations and design philosophy used in the TEAs. Figure 7b presents the capture cost comparisons with detailed breakdowns where similar values are obtained for each item considered in the annualized CapEx, fixed O&M, and variable O&M including steam and electricity costs. The final capture cost in this work was found to be 109.33 €/t_{CO_2} which is similar to the capture cost reported by Nwaoha et al. (2.41% difference). A detailed breakdown of each equipment cost and utility cost presented in Figure 7 can be found in Appendix A, Figure A1, Figure A2, and Table A6.

5.2 Overall performances of the MEA process at varying flue gas flow rates and CO_2 concentrations

After the simulation and TEC validation, the MEA process was studied at various flue gas flow rates and CO_2 inlet concentrations as presented in section 4. At a fixed capture scale, the absorber column is the main part to be significantly impacted by an increase in the flue gas CO_2 concentration. This is due to the fact that when CO_2 concentration rises at a fixed capture scale, the volumetric flow rate of flue gas decreases but the remaining parts of the capture plant will only be minimally impacted since the amount of CO_2 provided to the capture plant is constant across all variations of concentrations. Therefore, a more detailed study on the absorber column design was performed across the concentration and flue gas flow rates focusing on the macro-scale evaluation of the MEA process.

In the literature, many authors investigated the influence of CO_2 concentrations on the MEA capture process. Gardjarsdóttir et al. (2015) varied the flue gas CO_2 concentration from 5 to 40 mol% while setting a constant solvent lean loading value of 0.25. Husebye et al. (2012) also assumed a constant lean loading value of 0.206 across the CO_2 concentration variations (2.5 – 50.5 mol%). In this study, sensitivity studies were performed at each considered CO_2 concentration to identify the corresponding optimum lean loading which gives the minimum specific reboiler duty. To identify the optimum lean loadings, three steps are involved in this study: (1) run the MEA Aspen Plus model with equilibrium mode to identify the minimum lean loading at selected CO_2 inlet concentrations; (2) At a fixed CO_2 concentration, the flue gas flow rate is varied to correspond to different capture scales and to identify the scale effects on the lean loadings; (3) The MEA model is switched to rate-based model to rigorously study the absorber considering mass transfer limitations and to identify optimum lean loadings. In the last step, the lean loadings obtained from the equilibrium runs were used as initial points to help find the rate-based lean loadings. An example of lean loading versus reboiler duty at CO_2 inlet concentrations of 11% and 30% is presented in Figure 8 where grey markers (triangles and circles) are MEA systems evaluated at 1 Mt/y capture scale (high flue gas flow rates) while the '×' and '+' markers (orange and blue points) are evaluated at the same inlet concentration of 11% and 30% respectively but at lower capture scale.

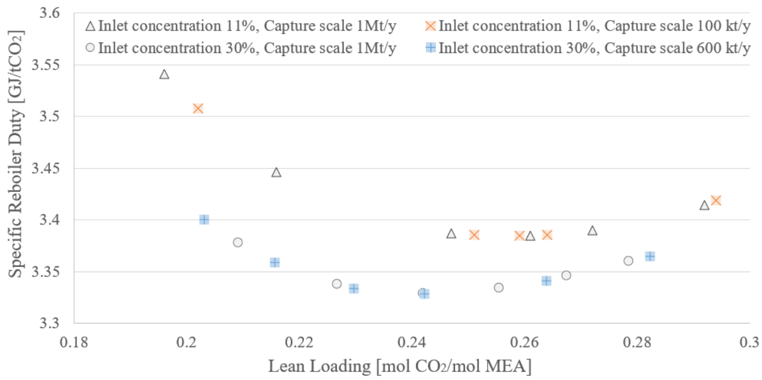

Figure 8. Specific reboiler duty and lean loading at different CO_2 concentrations and different capture scales

The shape of the curve is due to the fact that the reboiler duty is composed of three elements, namely: (1) heat needed to reach the regeneration temperature, (2) heat from CO_2 desorption that remains constant at a fixed capture rate, (3) heat provided to the reboiler to generate steam for regeneration. On the left-hand side of Figure 8, less solvent flow rate is required but a higher reboiler duty is needed to reach the desired level of solvent regeneration and corresponding low lean loading. As the lean loading increases, the solvent flow rate in the solvent loop increases and the system needs more energy to heat up the water present in the solvent and therefore, the required reboiler duty increases. This leads to an optimum lean loading (corresponding to an optimum solvent flow rate) and Figure 8 shows that this optimum is shifted to the left as the CO_2 inlet concentration increases from 11 to 30%. Another observation is that the specific reboiler duty profiles do not depend on the capture scale (flue gas flow rates) at fixed inlet CO_2

concentration. Flue gas flow rates were varied between 40.3 – 161.4 × 10^3 Nm³/h (0.5 – 2 kmol/s) to study the impact of the process scale on the process optimum operating points and the results are reported in Figure 9.

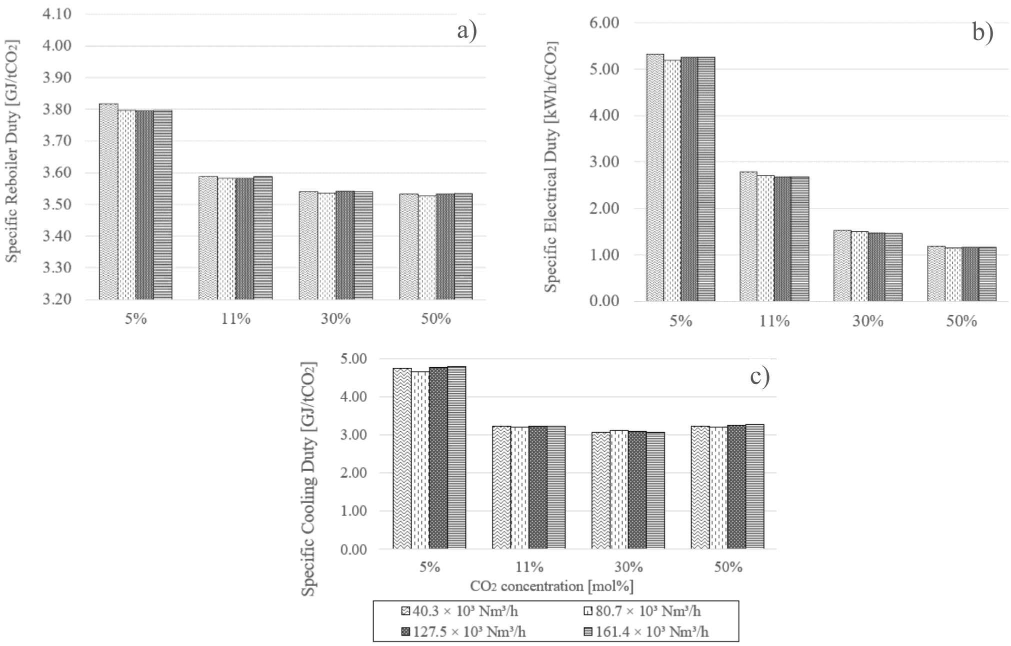

Figure 9. Specific energy consumption profiles across the CO_2 inlet concentrations: a) specific reboiler duty [GJ/tCO_2], b) specific electrical duty [kWh/tCO_2], c) specific cooling duty [GJ/tCO_2]

At a fixed CO_2 inlet concentration as shown in Figure 9, the specific energy requirements are observed to be independent of the flue gas flow rates (or capture scales) with variations lower than 5%, but to vary depending on CO_2 inlet concentrations. It is observed that the specific reboiler and electrical duties rapidly decrease and then level off as CO_2 concentration increased. However, the specific cooling duty also decreases rapidly as the CO_2 concentration increases from 5 to 30% but then very slightly increases at 50% concentration. This is due to the fact that absorption is an exothermic reaction, and more heat needs to be evacuated with high concentrations of CO_2 in the flue gas. As an overall trend, the influence of the CO_2 concentration at a fixed capture scale of 1 Mt/y and a capture rate of 90% is reported in Table 12. The gas flow rate is set by the capture scale and CO_2 concentration, while the lean solvent flow rate (L) is varied to achieve minimum reboiler duty.

Table 12. L/G, optimum lean loadings and specific reboiler duty across the selected CO_2 inlet concentrations at a capture scale of 1 Mt/y

CO_2 concentration [%]	L [t/h]	G [t/h]	L/G [kg/kg]	Lean loading	Rich loading	Reboiler Duty [GJ/t]
5	2247.2	1555.4	1.4	0.260	0.493	3.81
11	2085.5	695.8	3.0	0.256	0.509	3.58
30	1945.5	297.2	6.5	0.246	0.516	3.54
50	1843.3	196.6	9.4	0.235	0.520	3.53

As can be seen in the trend, the range of the flue gas CO_2 concentration (5 – 50 mol%) strongly influences process performances. The specific reboiler duty decreases exponentially with increasing CO_2 concentration while reaching a plateau-like trend at higher CO_2 concentration. From Table 12, it appears that to achieve the minimum reboiler duty, both the L/G and the loading intervals need to increase at high CO_2 concentrations. However, the total solvent flow rate decreases, although not as much as the gas flow rate, hence the larger L/G ratio. As an overall trend, to achieve a specific capture rate at a given capture scale, either solvent flow rates can change or more energy can be provided to the stripper to achieve leaner solvents (lower lean loading), which increases solvent capacity (defined as the difference between lean and rich loadings).

Another important design parameter considered during the data generation was absorber dimensions. For example, at a capture scale of 1 Mt/y and a given inlet CO_2 concentration, the diameter of the absorber was estimated based on the flue gas volumetric flow rate. The cost optimum height was then identified through a sensitivity analysis aiming at minimizing the total hourly cost [k€/h] of the column. This hourly cost corresponds to the TAC divided by the annual operating hours, representing the average cost of operation per hour. At the capture scale of 1 Mt/y, the cost optimum size of the column (m^3) decreases as the CO_2 inlet concentration increases as shown in Figure 10.

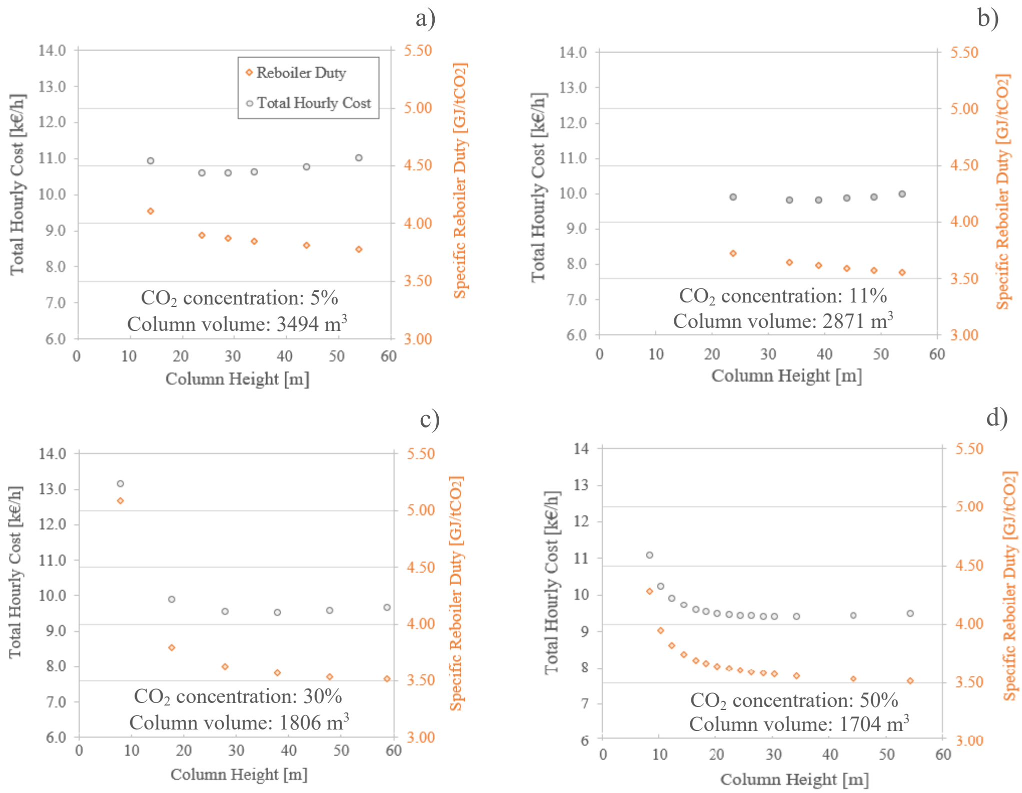

Figure 10. Influence of column height on the reboiler duty [GJ/t_{CO_2}] and the total hourly cost [k€/h]: cases at a fixed capture scale of 1 Mt/y at a) 5%, b) 11%, c) 30% and d) 50% CO_2 inlet concentrations

For each concentration considered in Figure 10, Table A5 in Appendix A presents the dimensions of each absorber column at a capture scale of 1 Mt/y. As an overall trend, an increase in the column height reduces the required reboiler duty as a taller column allows longer contacting time between the flue gas and the solvent which reduces the need for a low lean loading. However, a continued increase in the column height will result in more CapEx which will eventually dominate the OpEx so that a trade-off appears to minimize the total hourly cost of the absorber.

As described in Figure 8, specific reboiler duties and the corresponding optimal loadings are independent of the capture scales. Therefore, after obtaining optimized dimension at e.g. 1Mt/y capture scale, technical parameters including the optimal lean and rich loadings (as presented in Table 12) were fixed at a given inlet CO_2 concentration while flue gas flow rates were varied to generate data points at different capture scales. The results of specific energy duties, column dimensions and the L/G ratios across the range of capture scales (31-1250 kt/y) are shown below.

Table 13. Technical parameters across the capture scales (31 - 1250 kt/y) at 30 mol% CO_2 inlet concentration

Gas Flow Rate [Nm^3/h]	Capture Scale [kt/y]	Column Height [m]	Diameter [m]	H/D	Lean Loading	Rich Loading	L/G	Reboiler Duty [GJ/t]	Elec. Duty [kWh/t]	Cooling Duty [GJ/t]
6 697	31.24	28.8	1.8	16.0	0.243	0.516	6.5	3.544	1.51	3.084
40 345	187.49	32.7	4.0	8.2	0.242	0.517	6.4	3.540	1.52	3.068
80 690	374.99	37.8	5.0	7.6	0.247	0.516	6.6	3.536	1.50	3.107
127 491	592.48	37.8	6.0	6.3	0.243	0.516	6.5	3.543	1.47	3.084
161 381	749.98	37.8	7.0	5.4	0.242	0.517	6.4	3.540	1.47	3.068
213 023	1000.00	37.8	7.8	4.8	0.246	0.516	6.5	3.543	1.48	3.107
270 716	1250.59	37.8	9.0	4.2	0.245	0.516	6.5	3.541	1.48	3.094

Each row of data in Table 13 is a result of the simulation run at various capture scales where corresponding flue gas flow rates were calculated at 30 mol% CO_2 inlet concentration (results obtained at other concentrations can be found in Table A2-A4 in Appendix A). As flue gas flow rates are varied at a fixed CO_2 inlet concentration, the total amount of captured CO_2 also varies. At each gas flow rate, the required lean solvent flow rate and diameter of the absorber were re-calculated using Eq. 7 and Eq. 8, and entered as a design input in the Aspen Plus model. As observed in the lean loading versus reboiler duty curves in Figure 8, the location of the optimum lean and rich loading is not influenced by the change in capture scales at a fixed CO_2 inlet concentration and therefore, the reference case's (1Mt/y case) lean and rich loadings were used throughout the different capture scales. Since the specific energy consumptions are intrinsic variables to the capture scale, at the same lean/ rich loadings, the height of the absorber was varied to achieve specific energy requirements close to (or equal to) the base case's specific consumption. Then, the L/G ratio was recorded and throughout the capture scale and it was observed that the L/G ratio stayed at a value close to 6.5 (the maximum relative deviation of 2.23% within the data points). The same approach was taken for other CO_2 concentrations/ flue gas flow rates, with simulated data points presented in the Appendix, table A1.

5.3 Correlation parameters and relative errors

One of the main goals of this study is to develop correlations describing technical and economical parameters of a typical amine carbon capture technology as a function of CO_2 molar concentration and flue gas flow rate. Such short-cut cost correlations provide a quantitative method to rapidly estimate TEC and specific energy duties at various capture scales and flue gas CO_2 contents from different industries. Eq. 11 and Eq. 12 are used to describe the aforementioned parameters and the fitting parameters with the boundaries are presented in Table 14.

Table 14. Estimated parameters for the MEA process.

α	β	γ	n	m	x_{co_2} [mol%]	F [10^3 Nm^3/h] [a]
			TEC [M€$_{2023}$] [b]			
2.1673	0.8092	-0.00332	0.5291	0.8391	$5 \leq x_{co_2} \leq 50$	$4.03 \leq F \leq 1\,613.81$
			Specific Cooling Duty $\left[\frac{GJ}{t_{co2}}\right]$			
10.040	2.905	—	-33.73	0.2108	$5 \leq x_{co_2} \leq 50$	$4.03 \leq F \leq 1\,613.81$
			Specific Reboiler Duty $\left[\frac{GJ}{t_{co2}}\right]$			
1.471	3.560	—	-35.83	-0.0158	$5 \leq x_{co_2} \leq 50$	$4.03 \leq F \leq 1\,613.81$
			Specific Electrical Duty $\left[\frac{kWh}{t_{co2}}\right]$			
10.420	2.164	—	-23.49	-1.2350	$5 \leq x_{co_2} \leq 50$	$4.03 \leq F \leq 1\,613.81$

[a] Although the full ranges for CO_2 concentration and flow rates are given in the present table, it is important to recall that for each concentration, the correlation is only valid within the annual capture scale of 31 - 1250 kt/y, as indicated by the ranges presented in Table A1.
[b] For TEC, $\alpha + (\beta \cdot {x_{co_2}}^{n} + \gamma) \cdot F^{m}$ format is used while for the specific duties, $\alpha \cdot e^{n x_{co_2}} + \beta \cdot e^{m x_{co_2}}$ format is used as presented in section 4.

For each correlation, average relative errors resulting from the comparison between correlation prediction and accurate calculation are presented in Table 15. As Hasan et al. (2012) pointed out, errors as high as 10 to 20% are common in the estimation of TEC when process scales are varied. Therefore, the average errors as well as the maximum relative errors presented in Table 15 justify these proposed correlation parameters.

Table 15. Average, minimum and maximum relative errors of the correlations for amine technology

MEA System Correlations	Average relative error [%]	Minimum relative error [%]	Maximum relative error [%]	Correlation coefficient, R^2
TEC	**4.9**	0.03	10.22	0.995
Steam	**0.1**	0.01	0.42	0.997
Electricity	**1.1**	0.01	4.06	0.999
Cooling	**0.5**	0.004	3.72	0.998

For the compression loop, regardless of the CO_2 inlet concentrations, as long as the capture scale is kept constant, constant specific TECs were observed. This is because the TEC of the compression loop is strongly related to the amount of CO_2 to be compressed to 150 bar. In terms of energy consumption, the specific electrical and cooling duties (per ton of captured CO_2) stayed the same across the capture scales as the process settings such as inlet/ outlet pressure of the compression loop and the cooling temperatures are kept the same throughout the data generation. In other words, the specific energy consumption for the compression loop is directly proportional to the capture scale. The trends can be found in Figure 11. The equation describing specific TEC [M€$_{2023}$/tCO_2/h] of the compression loop in terms of capture scale (S) [t/h] is presented in Eq. 12 below.

$$\mathrm{TEC}_{\,compression} = 0.3334(S)^{-0.503} \qquad \text{Eq. (12)}$$

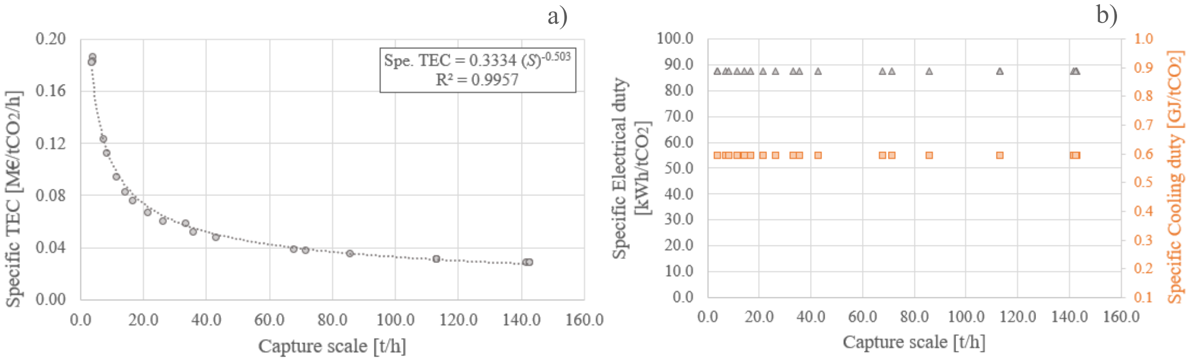

Figure 11. a) Specific TEC and b) specific energy consumption of compression loop

5.4 *Case studies*

In this section, the correlations developed in this study are applied to several case studies representing various industrial flue gas conditions at 1 bar. For each case, the correlations were used to estimate the TEC and energy consumptions based on case-specific parameters such as CO_2 inlet concentration, flue gas flow rate (or annual capture scale), assumptions, CapEx estimation methods, boundaries, utility costs, etc. Subsequently, the results were used to calculate the CO_2 capture cost, and the differences in CapEx, OpEx and overall capture costs were analysed and compared. A detailed breakdown of the results is presented in Appendix D for all case studies.

5.4.1 Case 1: Cement industry at 11.5% CO_2 inlet concentration

The first case study is to estimate the carbon capture cost of the benchmark case study used in this paper (Nwaoha et al., 2018) where elements used in the capture costs are shown below in Table 16 with assumptions and estimation methods used in the analysis. The correlations developed in this paper are used to estimate the TEC at a given CO_2 concentration and flue gas flow rate. The DOE/NETL method as presented in the work of Nwaoha et al. (2018) was then applied to the TEC to obtain the CapEx while the case-specific utility costs presented in Table 10 are used with the energy consumptions predicted using the correlations to estimate OpEx. Since the costing data for correlations are assumed to be based in the US Gulf, a location factor of 1.10 (in 2018) for Canada is used for the CapEx-related items to account for the location differences. For labour-related items, a labour location factor of 1.82 (in 2018) was used (IEAGHG, 2018).

The scope of this case study includes the CO_2 capture process and compression loop. It is worth noting that the compression loop simulated in this study involves 3 compressors and a pump (section 3.2), which is different from the system presented in the work of Nwaoha et al. (2018) which consists of 4 compressors. Therefore, there may be some differences in the compression costs. However, the cost results in Nwaoha et al. (2018) present the sum of the CO_2 capture process and compression loop without the breakdowns. Therefore, it was not possible to decouple the capture and compression costs separately. Nonetheless, the costs of carbon capture, compressions, and the total costs (the sum of the capture and compression costs) obtained from the correlations are presented below.

Table 16. Carbon capture cost results using correlations versus literature results (Nwaoha et al., 2018)

	This work (Correlation)	**Nwaoha et al. (2018)**	**Dif. [%]**
Technology	MEA 30 wt%	MEA 30 wt%	
Costing based year	2023	2023	
Equipment costing methods	Nwaoha's Vendor/OEM	Nwaoha's Vendor/OEM	
CapEx costing method	DOE/NETL	DOE/NETL	
Location	Quebec, Canada	Quebec, Canada	
Capture Scale [Mt/y]	0.7	0.7	
CO_2 concentration [mol%]	11.5	11.5	
Life time [y]	30	30	
Scope	CO_2 Capture, compression	CO_2 capture, compression	
CapEx [M€]	224.1	177.6	+20.76
OpEx [M€/y]	56.13	61.90	-10.27
Specific OpEx [€/t]	80.19	88.43	
Specific CapEx [€/t/y]	320.2	253.7	
CO_2 capture cost [€/t]	102.78	NA	
Compression cost [€/t]	7.20	NA	
Total cost (sum of above) [€/t]	**109.98**	**112.03***	**-1.86**
% of Ann. CapEx in capture cost	27%	21%	
% of OpEx in capture cost	73%	79%	

*The capture cost reported in the work of Nwaoha et al. (2018) consists of both CO_2 capture and compression steps without breakdowns of each step.

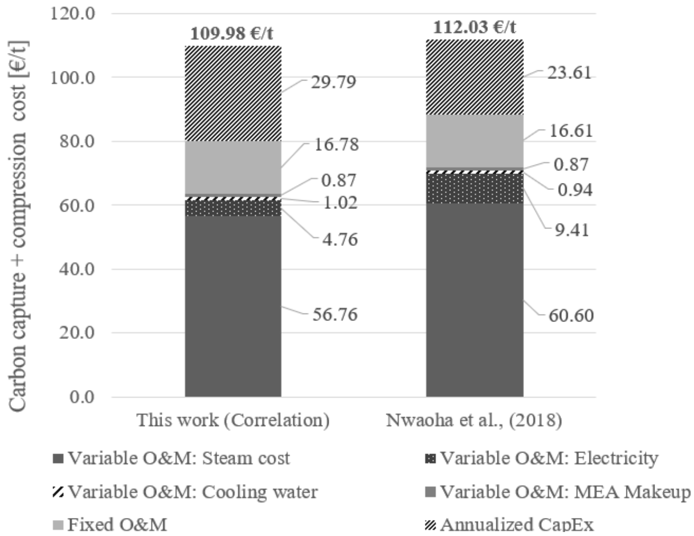

Figure 12. Detailed breakdown of CO_2 capture + compression cost [€/t]: case 1

In terms of the CapEx, the difference is +20.76% while the difference observed in the OpEx is -9.05%. It is emphasized that during the data generation step explained in section 5.2, the absorber column height was optimized in terms of TAC and resulted in higher columns which means heavier column weights. In addition, the column weight calculated from the Aspen Plus was used during the data generation which was more detailed than the heuristics used in Nwaoha's work and therefore, the results from the simulation are believed to reflect more realistic column weights and as a result, a higher CapEx is observed using correlations. The difference in CapEx is within the uncertainty proposed by AACE in Table 2 and the CapEx is annualized over the lifetime of 30 years with a discount rate of 8.5% to calculate the capture cost. Therefore, the difference in CapEx has less impact on the final capture costs than the difference found in OpEx (Figure 12). Within the specified system boundary of CO_2 capture and compression, the OpEx heavily influences the final capture cost where the percentage of the OpEx in the final capture cost is more than 70%. This indicates that using case-specific utility costs is very important when estimating capture costs. Simply comparing the final capture cost alone is thus not enough to establish a fair comparison between studies although this is a common practice in the literature today.

5.4.2 Case 2: Cement industry at 22% CO_2 inlet concentration

The correlations were applied to a system with a higher CO_2 inlet concentration of 22 mol%. The case-specific utility costs from the work of Roussanaly et al. (2018) were used, i.e. 9.33 €/GJ of LP steam, and electricity cost of 0.082 €/kWh. A lifetime of 25 years and a discount rate of 8% are used in this case study. The comparison of the correlation results and the literature is shown in Table 17 and Figure 13.

Table 17. Carbon capture cost results using correlations versus literature results (Roussanaly et al., 2018)

	This work (Correlation)	**Roussanaly et al. (2018)**	**Dif. [%]**
Technology	MEA 30 wt%	MEA 30 wt%	
Costing based year	2023	2023	
Equipment costing methods	Nwaoha's Vendor/OEM	Bottom-up	
CapEx costing method	Roussanaly et al. (2018)	Roussanaly et al. (2018)	
Location	Euro region	Euro region	
Capture Scale [Mt/y]	0.76	0.76	
CO_2 concentration [mol%]	22.0	22.0	
Life time [y]	25	25	
Scope	CO_2 Capture, compression	CO_2 capture, compression	
CapEx [M€]	141.77	175.96	-24.1
OpEx [M€/y]	43.40	45.55	-4.9
Specific OpEx [€/t]	57.10	59.93	
Specific CapEx [€/t/y]	186.54	231.53	
CO_2 capture cost [€/t]	64.47	NA	
Compression cost [€/t]	10.11	NA	
Total cost (sum of above) [€/t]	**74.58**	**81.62***	**-9.4**

% of Ann. CapEx in capture cost	23%	27%
% of OpEx in capture cost	77%	73%

*The capture cost reported in the work of Roussanaly et al. (2018) consists of both CO_2 capture and compression steps without breakdowns of each step.

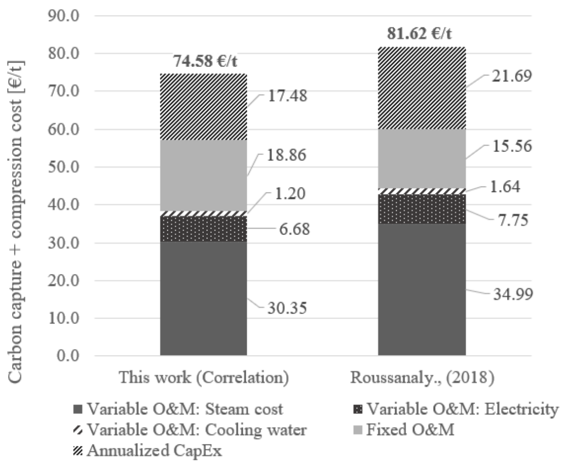

Figure 13. Detailed breakdown of CO_2 capture + compression cost [€/t]: case 2

Note that Roussanaly et al. (2018) used a bottom-up method to estimate equipment cost but without giving any further specification about the equipment costing. Regarding the CapEx costing method, they presented the main assumptions used to estimate the final CapEx. For example, from the total direct cost (TDC) which consists of the sum of the TEC and supporting facilities & total labour cost, a factor of 1.14 was applied to obtain the engineering, procurement and construction (EPC) cost. Then another factor of 1.19 was applied to the EPC to estimate the total plant cost (TPC). Then, 15% of TPC was assumed to be the owner's cost. Finally, the initial amine cost and the owner's cost were added to the TPC to obtain the final CapEx. Further details of this cost estimation method are provided in Appendix D. The same method was applied to the correlation results to establish a fair comparison. The location factor of 1.01 (in 2018) for Euro region is used for the CapEx-related items to account for the location differences in the correlation results which are based in the US Gulf. For the labour-related items, labour location factor of 1.7 (in 2018) was used (IEAGHG, 2018). The operating hour of 8003 h/y was used in the case study. There are some minor differences in the final compression pressure where 110 bar was used in Roussanaly et al. (2018) while 150 is implemented in this study. However, a detailed breakdown of compression cost is not available in the literature and therefore, the sum of the CO_2 capture and compression cost obtained from Roussanaly et al. (2018) is used in this case study. Using the correlations developed in this study, the capture process and compression system can be evaluated individually, as shown in Table 18 and providing detailed breakdowns are important practice to minimize ambiguity and errors in carbon capture costs. Overall, the correlation results show good predictions (9.4% difference) close to the literature values presented by Roussanaly et al. (2018) using the CapEx method from the benchmark literature while the capture cost of 77.39 €/t_{CO_2} (5.18 % difference) can be expected when DOE/NETL method was applied. The correlations developed in this work can flexibly adopt different assumptions and estimation methods as demonstrated above.

5.4.3 Case 3: Refinery at 8.2% CO_2 inlet concentration

In this case study, a system with a lower CO_2 concentration and a small capture scale of 0.3 Mt/y was studied following a report from the Norwegian research centre SINTEF (Sannan et al., 2017). Given the natural gas cost in the report (8.66 €$_{2023}$/GJ), a LP steam cost of 11.26 €$_{2023}$/GJ was computed following heuristics presented in the technical report of DOE (2003), where a factor of 1.3 was applied to the natural gas cost. For electricity, an estimation equation for the onsite power generation was considered (Ulrich and Vasudevan, 2006) based on the fuel cost resulting in a cost of 0.204 €$_{2023}$/kWh. Note that the utility costs used in this case study are estimations made by the current authors since these data were not available in the literature of Sannan et al. (2017). The equipment costs were prepared by SINTEF and the authors of the report claimed that the CapEx costing method presented in IEAGHG (2017) was used. The same CapEx method and assumptions (refer to Appendix D) presented in Sannan et al. (2017) are applied to the correlations developed in this work. A location factor of 1.01 (in 2018) for Euro region is used for the CapEx-related items to account for the location differences. The number of operating hours of 8400 h/y is used in the case study as described in the literature. The plant life-time of 25 years and a discount rate of 7.12% was used to annualize the CapEx. A comparison result is presented in Table 18.

Table 18. Carbon capture cost results using correlations versus literature results (Sannan et al., 2017)

	This work (Correlation)	**SINTEF: Sannan et al. (2017)**	**Dif. [%]**
Technology	MEA 30 wt%	MEA 30 wt%	
Costing based year	2023	2023	
Equipment costing methods	Nwaoha's Vendor/OEM	SINTEF	
CapEx costing method	IEAGHG, (2017)	IEAGHG, (2017)	
Location	Euro region (Netherlands)	Euro region (Netherlands)	
Capture Scale [Mt/y]	0.30	0.30	
CO_2 concentration [mol%]	8.20	8.20	
Life time [y]	25	25	
Scope	CO_2 Capture, compression, retrofitting (interconnecting)	CO_2 capture, compression, retrofitting (interconnecting)	
CapEx [M€]	157.57	194.49	-23.4%
OpEx [M€/y]	20.85	22.34	-7.2%
Specific OpEx [€/t]	69.03	73.97	
Specific CapEx [€/t/y]	521.76	644.01	
CO_2 capture cost [€/t]	68.98	NA	
Interconnecting cost [€/t]	30.99	30.99	
Compression cost [€/t]	14.31	NA	
Total cost (sum of above) [€/t]	**114.28**	**129.82***	-13.60%
% of Ann. CapEx in capture cost	40%	43%	
% of OpEx in capture cost	60%	57%	

*The capture cost reported in the work of Sanann et al. (2017) consists of CO_2 capture, interconnecting and compression costs. CapEx of each step is available however, OpEx is given as a lump sum of the three steps without breakdowns and therefore, it was not possible to decouple the costs. CW = Cooling Water

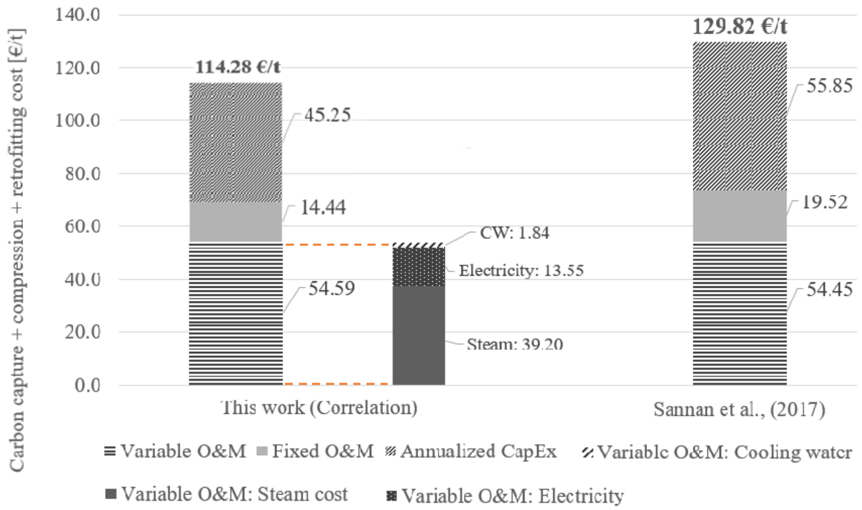

Figure 14. Detailed breakdown of CO_2 capture + compression + retrofitting cost [€/t]: case 3

The scope of the TEA includes CO_2 capture, compression, and retrofitting which involves interconnecting the existing plant to the capture plant. In order to allow for a proper comparison of the capture and compression costs between this study and Sannan et al. (2017), the retrofitting costs (both CapEx and fixed O&M) presented in the SINTEF report were added to the estimations obtained from the correlations. A breakdown of the capture cost is presented in Figure 14. For the results from the correlations, a breakdown of the variable O&M is presented in the exploded bar graph for each utility consumption. However, for the results from Sannan et al. (2017), full details of the variable O&M were not presented in the report except for the steam consumption and therefore, in this case, a lump sum is presented. The authors mention that 69% of natural gas consumption is linked to steam production for regeneration of MEA solvents. Based on this information, a steam cost of 37.57 €/t_{CO_2} was calculated, which is similar to the result obtained from the correlations (39.20 €/t_{CO_2} in Figure 15 with a difference of 4.2%). Overall, the results from the correlation approach provide a good estimation of the MEA system presented in the SINTEF report. It is stressed that the correlations only cover major elements of the MEA process and therefore, depending on the applications and/or literature, the depth of the required details may vary. It is advised to have a comprehensive understanding of the scope and elements considered in a TEA.

5.4.4 Gap between the literature and the reality

In the previous sections, the correlations developed in this paper were applied in various case studies to show correlations' flexibility and applicability to different cases. However, most of the literature, and reports available in the public domain are theoretical case studies and real-life examples are scarce today. There are few large-scale carbon capture plants in the world and the Petra Nova plant in Texas (Armpriester, 2017) is one of the first-of-a-kind (FOAK) amine-based carbon capture plants.

One of the big challenges in the carbon capture TEA literature is that there exists a gap between the results from the literature and the real-life plants and therefore, performing detailed validations on the TEA studies is not always straightforward due to the lack of available data and discrepancies in the costing methodologies. The lack of such data for real-life capture plants also causes many carbon capture TEA papers to rely on literature-based cost estimates and reduce the validity of the results compared to reality.

In the case of Petra Nova plant, the capture cost of 70 $€_{2020}/t_{CO_2}$ is reported in the literature (Global CCS Institute, 2021). However, details of the CapEx and OpEx are not fully available in the public domain and therefore, estimates provided in the DOE/NETL report (Armpriester, 2017) are used in this study for comparison purposes. It is emphasized that due to the lack of detailed data, it is very tricky to identify the actual costs associated with the system boundary considered in this study (carbon capture and compression only). From the given estimates in the report, the CapEx cost of the carbon capture unit and CO_2 compression/ dehydration was found to be 535.22 $M€_{2023}$. This appears to be 60% higher than the predicted value from the correlation developed in this study which is 334.40 $M€_{2023}$. One reason to explain the gap may be related to the fact that the costing methodologies used in the literature and in this work do not account for overprices related to a FOAK unit. In order to evaluate this assumption, a N^{th}-of-a-kind (NOAK) analysis was performed with a learning rate of 0.19 as a typical learning rate to see the cost gap between the literature and the reality and the results are shown in Figure 15.

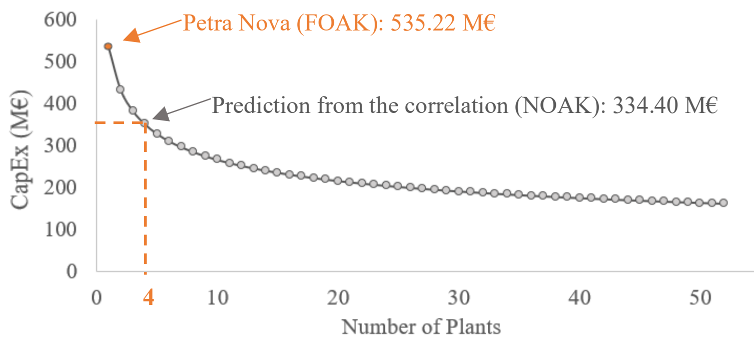

Figure 15. NOAK Analysis of the Petra Nova CapEx

From the NOAK analysis, it can be observed that after constructing 4 amine plants (indicated with dash lines in Figure 16), the cost of the capture unit can be lowered to 334 $M€_{2023}$. According to the CCS Knowledge (2018), who conducted a feasibility study of the Shand plant (which is their 2^{nd} carbon capture facility planned to be constructed) and compared it to the cost of the Boundary Dam carbon capture plant, a 67% capital cost reduction is expected due to the lessons learned from the previous plant, modularization and simplifications of the process. The percentage suggested by the CCS Knowledge may seem too optimistic since the usual learning rate is around 20%. Nevertheless, this N^{th}-of-a-kind approach can definitely contribute to explain the large gap observed between costing methodologies and industrial reality. Even if large contingency fees were used from the DOE/NETL costing methodology (see section 2.2), a significant gap can still be observed in Figure 16 and the only way to achieve cost estimates closer to reality would be to build more capture facilities and sharing the costing details.

Finally, Figure 16 also shows that industrial carbon capture costs are still quite high and that the learning-by-doing approach can be effective to achieve a significant cost reduction that could contribute to technological deployment. The usage of short-cut correlations developed in this work is envisioned to assist in such situations where rapid decisions are required for pre-feasibility studies to speed up the deployment rates. Nevertheless, based on the information above, the gap between the correlation results (or theoretical cost estimates) and reality can be explained in a reasonable manner, but comprehensive studies with full cost details (and data) are crucial in identifying the exact reasons for the differences.

5.4.5 Carbon capture cost profile

A last result that can be presented to evidence the simplicity of using the correlations is to evaluate the cost of the amine carbon capture technology for feed gas CO_2 compositions from 5 to 50% and flue gas flow rates of 40-161 × 10^3 Nm^3/h. These ranges were arbitrarily selected to show the overall capture cost profile presented in Figure 16. To obtain the OpEx, utility costs from Table 11 were used. Different markers are used to indicate different flue gas flow rates.

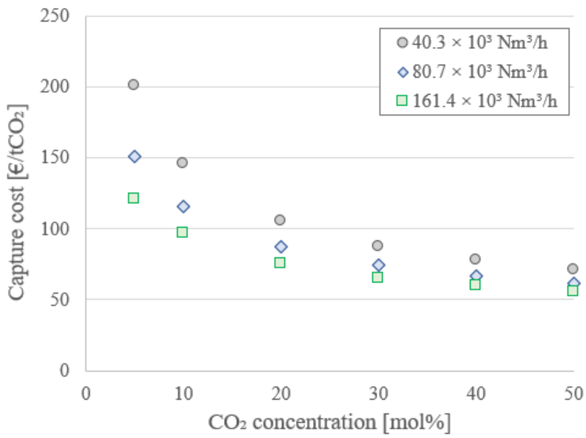

Figure 16. Capture cost for the MEA-based technology (cost includes carbon capture and compression)

The effect of the CO_2 inlet concentration is evident and is due to the concentrated inlet flue gas stream having a lower volume leading to lower equipment costs, but also to more efficient absorption. In terms of the flue gas flow rates, specific capture costs decrease as flow rates increase due to economies of scale. However, this effect of scale is relatively small due to the large share of the OpEx in the total cost of CO_2 capture.

Furthermore, these effects can be easily quantified at different capture scales and concentrations using the correlations presented in this work. The trends presented in Figure 16 will be even more insightful when comparing amine capture with other CC technologies such as physical absorption and membrane carbon capture options to identify the best technology depending on each industrial sector and flue gas conditions.

6. Conclusion and perspectives

The benchmark CO_2 capture technology, the MEA process was studied in detail and modelled at various CO_2 inlet concentrations and flue gas flow rates. A capture rate of 90% was imposed, along with compression loops to 150 bar. The MEA absorber in each selected operating condition was optimized in terms of TAC using a rate-based Aspen Plus model. Then, generated data points were used to develop correlations describing the total equipment cost (TEC) and specific energy consumption (steam, electricity, and cooling water requirements per ton of captured CO_2) depending on the capture scale and CO_2 concentration in the inlet gas. A comparison of costing methodologies was presented to show the current challenges in the TEA literature as well. The goal of this work is not to propose a new global guideline for CapEx and OpEx but to identify existing methodologies and to present flexible correlations for identifying the TEC and specific energy requirements as intermediate steps for the users to apply their own costing assumptions/ data to estimate capture costs.

One of the novelties of the correlations presented in this work is that different CapEx estimation methods can be applied based on the calculated TEC. With that, users willing to compare different CO_2 capture case studies (e.g. assuming MEA-based capture for different industrial sectors) are ensured to use consistent TEA assumptions which they can select and thus to reduce inherent structural errors arising from differences found in the existing CapEx estimates in the literature. In addition, case-specific utility costs can be applied to the specific energy consumption correlations to fine-tune scenarios where previously existing correlations in the literature were limited to estimate only the final OpEx without any flexibility in selecting utility costs.

Throughout several case studies, the applicability of the correlations across various concentrations and flue gas flow rates is presented, allowing for a validation of the proposed correlations. Also, when validating the results, it is a good practice to show a detailed breakdown of capture costs instead of solely comparing the total capture costs as there are many elements involved in the calculation of the capture costs. Lastly, the correlations can serve as a shortcut model to quickly estimate the capture cost of the MEA process for different types of flue gases (flow rates and CO_2 concentrations) while ensuring the validity of the results which were successfully compared with literature results. This advantage of the correlation approach in this paper could be helpful as a guideline in the decision-making processes for carbon capture deployments.

In the context of this paper, due to the computational challenges with the MEA model, a limited number of data points were arbitrarily selected to develop the correlations with the aim to cover a broader range, with CO_2 concentrations ranging from 5 to 50 mol% and capture scales from 31 to 1250 Mt/y. It would be beneficial to apply data sampling techniques such as the Latin Hypercube Sampling (LHS) thoroughly to refine and gain deeper insights into the correlations, possibly to study the influence of other flue gas variables as well (e.g. in case flue gas may be available at varying pressure). Also, the study can be conducted at various capture rates other than the ubiquitous capture rate of 90% in the literature since higher capture rates can be beneficial with a low additional cost. The CO_2 emissions from the MEA process can also be added to the correlation outputs to estimate the avoidance cost to accurately represent the performance of the capture technology, although this would require additional assumptions regarding the type of available energy supply for the CO_2 capture process.

In addition, no heat integration was performed in this paper but including such systems as a form of credits to the carbon capture costs would provide more meaningful metrics. This approach could involve considering an advantageous cost for LP steam. Also, generally it was observed that the retrofitting and interconnecting costs as well as pre-treatments or further post-treatments are not studied in detail in the literature. Adding these extra costs on top of the CO_2 capture and compression costs would provide a more realistic estimation of the implementation of capture technologies. Lastly, the same techniques used to develop the MEA correlations will be applied to other capture technologies such as physical absorption and membrane options to enrich the comparison studies using the correlations. The final objective of this work is thus to develop a decision support tool to provide easy comparison of different CO_2 capture technologies for varying industrial flue gas conditions. This will allow industrial users to rapidly assess the benefits and costs of CO_2 capture for their local conditions and specific emissions, leading to a better identification of the optimum technologies for successful CO_2 capture deployment.

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Appendix A

Table A1. Data generated for correlation developments

CO_2 concentration [mol%]	Gas flow rate [Nm^3/h]	Capture scale [kt/y]
5	40 345	31
	80 690	62
	127 491	99
	161 381	125
	242 071	188
	484 142	375
	1 274 908	990
	1 613 808	1250
11	18 559	31
	40 345	73
	80 690	146
	104 090	188
	127 491	230
	161 381	291
	550 309	992
	734 283	1250
13	45 187	91
	90 373	182
	180 747	364
21	60 251	196
23	51 363	183
30	6 697	31
	40 345	187
	80 690	375
	127 491	592
	161 381	750
	213 023	990
	270 716	1250
31	11 046	53
36	12 014	67
	36 043	201
47	13 717	100
	27 435	200
	54 870	400
	109 739	799
50	4 035	31
	20 173	156
	40 345	312
	80 690	625
	127 491	990
	161 381	1250

Table A2. Technical parameters across the capture scales (31 - 1250 kt/y) at 5 mol% CO_2 inlet concentration

Flow rate [Nm^3/h]	Capture scale [kt/y]	Packed Height [m]	Diameter [m]	H/D	Lean Loading	Rich Loading	L/G	Reboiler Duty [GJ/t]	Elec. Duty [kWh/t]	Cooling Duty [GJ/t]
40 345	31.25	23.8	3.00	7.93	0.2607	0.4838	1.5	3.82	5.32	4.76
80 690	62.50	23.8	3.69	6.4	0.2607	0.4859	1.5	3.80	5.19	4.66
127 491	98.75	23.8	5.00	4.8	0.2535	0.4864	1.5	3.80	5.25	4.78
161 381	125.00	23.8	5.22	4.6	0.2599	0.4857	1.5	3.80	5.25	4.79
1 274 908	989.98	23.8	13.59	1.8	0.2599	0.4847	1.5	3.81	5.17	4.81
1 613 808	1249.96	23.8	16.00	1.5	0.2599	0.4855	1.5	3.80	5.21	4.82

Table A3. Technical parameters across the capture scales (31 - 1250 kt/y) at 11 mol% CO_2 inlet concentration

Flow rate [Nm^3/h]	Capture scale [kt/y]	Packed Height [m]	Diameter [m]	H/D	Lean Loading	Rich Loading	L/G	Reboiler Duty [GJ/t]	Elec. Duty [kWh/t]	Cooling Duty [GJ/t]
18 559	33.01	28.8	2.40	12.0	0.252	0.510	2.9	3.59	2.67	3.20
40 345	72.87	30.8	2.84	10.8	0.257	0.508	3.0	3.59	2.79	3.23
80 690	145.73	31.8	4.04	7.9	0.254	0.509	3.0	3.58	2.71	3.22
127 491	230.37	31.8	5.11	6.2	0.256	0.509	3.0	3.58	2.69	3.22
161 381	290.87	31.8	5.49	5.8	0.256	0.508	3.0	3.59	2.68	3.23
550 309	991.91	33.8	10.22	3.3	0.256	0.509	3.0	3.58	2.64	3.22
734 283	1250.12	33.8	12.00	2.8	0.257	0.509	3.0	3.57	2.64	3.22

Table A4. Technical parameters across the capture scales (31 - 1250 kt/y) at 50 mol% CO_2 inlet concentration

Flow rate [Nm^3/h]	Capture scale [kt/y]	Packed Height [m]	Diameter [m]	H/D	Lean Loading	Rich Loading	L/G	Reboiler Duty [GJ/t]	Elec. Duty [kWh/t]	Cooling Duty [GJ/t]
4 035	31.23	41.8	2.00	20.9	0.2312	0.5212	9.2	3.53	1.19	3.21
40 345	312.49	44.3	4.00	11.1	0.2318	0.5207	9.2	3.53	1.18	3.23
80 690	624.98	44.3	6.00	7.4	0.2323	0.5213	9.2	3.53	1.16	3.22
127 491	989.98	44.3	7.00	6.3	0.2353	0.5200	9.4	3.53	1.17	3.25
161 381	1249.96	44.3	7.80	5.7	0.2353	0.5196	9.4	3.53	1.17	3.26

Table A5. Cost optimum absorber dimensions at different CO_2 concentrations

	Unit	CO_2 inlet concentration			
		5%	11%	30%	50%
Diameter	[m]	13.7	10.4	7.8	7
Column height	[m]	23.7	33.8	37.8	44.3

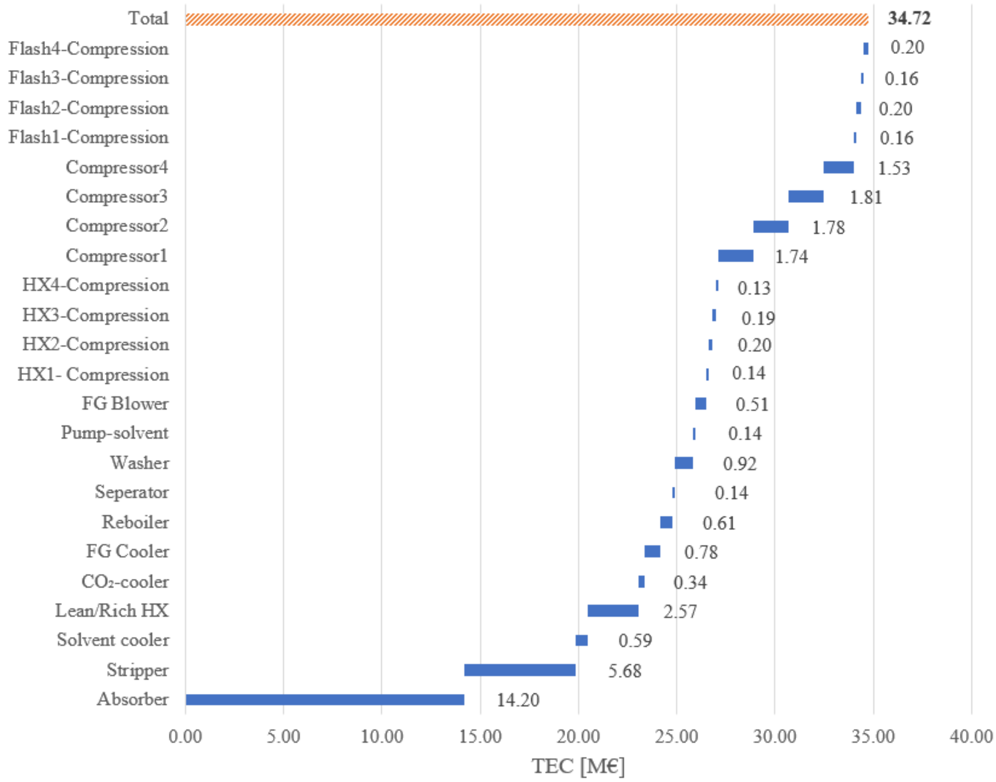

Figure A1. Detailed breakdown of the equipment cost used in the benchmark case study

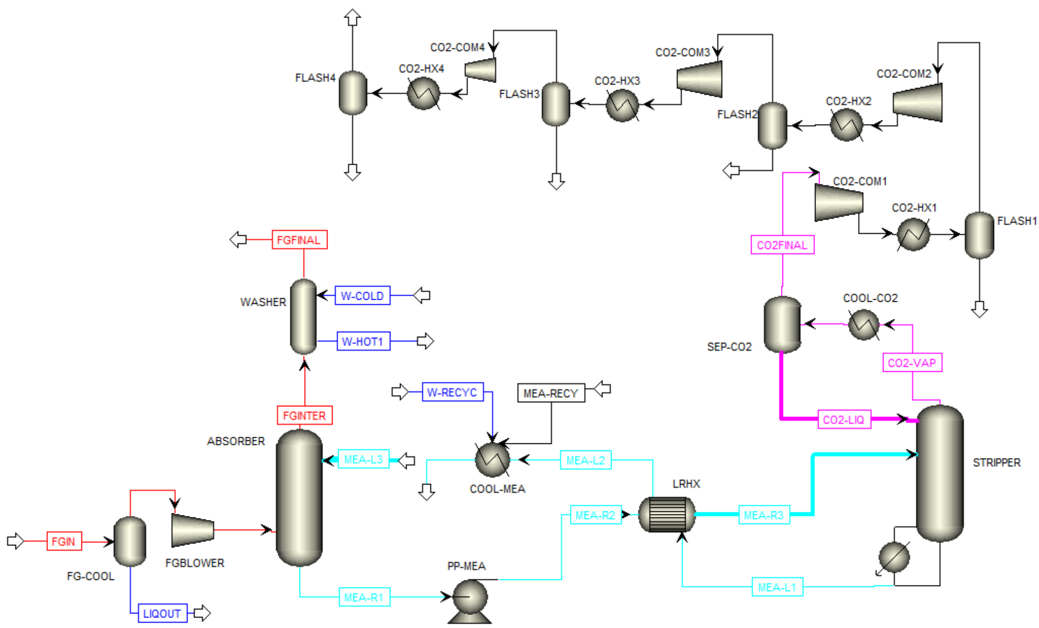

Figure A2. Schematic diagram of the MEA CO_2 capture process in Aspen Plus

Table A6. Summary of the utility consumptions and costs used in the benchmark case study

Cooling	**GJ/y**	**M€/y**	**€/t**
FG cooler	504092.86	0.14	0.19
CO_2 cooler	646910.62	0.17	0.24
Solvent cooler	1180383.90	0.32	0.46
Compression loop	389664.27	0.11	0.16
Total (Sum of above)	3033487.50	0.82	**1.05**
Electricity	**kWh**	**M€/y**	**€/t**
Blower	10164419.50	0.57	0.81
Pump-solvent	764179.31	0.04	0.06
Compression loop	57465639.97	5.10	7.29
Total (Sum of above)		5.71	**8.16**
Steam (LP)	**GJ/y**	**M€/y**	**€/t**
Reboiler	2112645.03	41.83	**59.76**

Appendix B: Column weight calculation

To estimate the weight of columns, following assumptions presented in Table B1 were applied.

Table B1. Assumptions used in the column weight calculation (Nwaoha et al., 2018)

Assumption	**value**
Density of stainless steel [kg/m^3]	7980
Wall thickness [m]	0.00965
Corrosion allowance [m]	0.00889

The absorber height (h) and the radius (r) used in the benchmark case study was 21.95 m and 5.04 m respectively. These dimensions were used to calculate the volume of the column:

$$V_{column} = \pi r^2 h \qquad \text{Eq. (B1)}$$

The obtained column volume was 1748.17 m^3 which is the volume of the inner column layer. To estimate the outer layer of the column accounting for the wall thickens and the corrosion allowance, an outer radius was estimated where the sum of the wall thickness and corrosion allowance of 0.0185 m was used.

$$r_{outer} = \frac{(2 \times r) + (2 \times 0.0185)}{2} = 5.053 \text{ m} \qquad \text{Eq. (B2)}$$

The volume of the outer layer was then estimated as 1761.07 m^3. Therefore, the volume of the shell was estimated by calculating the difference between the volumes: 1761.07 m^3 - 1748.17 m^3= 12.9 m^3. Then, the column weight (W_{column}) was calculated by multiplying the density of stainless steel:

$$W_{column} = 12.9 \text{ m}^3 \times \frac{7980 \text{ kg}}{m^3} = 102937.13 \text{ kg}$$

Appendix C: Supplementary information on the absorber sizing via parametric optimization

The reboiler duty is affected by several process parameters, including column height, and solvent flow rates. To identify optimal values, a sensitivity analysis was conducted on each of parameter individually, with other factors (the flue gas flow rate, capture rate and absorber column diameter) held constant. Firstly, the height of the absorber was adjusted to evaluate its effects on specific reboiler duty and the TAC (Total Annual Cost) of the column at an initial solvent flow rate. An increase in absorber height reduces the reboiler duty, as it provides more contact time between the solvent and incoming gas. The capture rate of the MEA Aspen Model was maintained at 90% using a design specification in Aspen Plus where the reboiler duty requirement in the stripper was adjusted to meet the capture rate. The following steps were taken in this study:

Step 1: **Identifying optimal lean loadings using equilibrium absorber model**

Using an equilibrium absorber column model with 8 stages in Aspen Plus, an optimum loading at 30% CO_2 concentration was identified to minimize the reboiler duty. The optimum lean loading was found to be around 0.242 and this was used as a guideline to obtain a feasible lean solvent flowrate in Step 2. Figure 8 presents an example of lean loading versus specific reboiler duty plot at 30% CO_2 concentration.

Step 2: Estimating lean solvent flow rate and column diameter

The lean solvent flow rate required to capture 90% of the CO_2 in the flue gas entering the absorber is estimated using Eq. (7) below:

$$L_{Lean} = \frac{G m_{co_2} \varphi_{co_2}}{100z(\alpha_{Rich} - \alpha_{Lean})} \cdot \left[\frac{M_{MEA}}{44.009} \left(1 + \frac{1 - \omega_{MEA}}{\omega_{MEA}} \right) + z\alpha_{Lean} \right] \quad \text{Eq. (7)}$$

Where G is the mass flow rate of the feed gas (kg/s), m_{co_2} is the mass fraction of CO_2 in the flue gas (%), φ_{co_2} is the capture rate (%), z is the number of molar equivalents/ mole of amine (1 for MEA), α_{Rich} and α_{Lean} are the lean and rich loading (mol CO_2/mol MEA) respectively as retrieved from Aspen Plus in Step 1, M_{MEA} is the molar weight of MEA (kg/kmol) and ω_{MEA} is the MEA weight percentage. As an example, the following inputs were used to estimate an initial lean solvent flowrate at 30 mol% CO_2 inlet concentration and 1 Mt/y capture scale case:

Table C1. Input variables for the estimation of initial solvent flowrate using Eq. (7)

Input variables	value
G [kg/s]	82.5
m_{co_2} [%]	30
φ_{co_2} [%]	90
z	1
α_{Rich} [mol CO_2/mol MEA]	0.543
α_{Lean} [mol CO_2/mol MEA]	0.242
M_{MEA} [kg/kmol]	61.08
ω_{MEA} [wt%]	30

Using Eq. (7) and the input variables listed in Table C1, a lean solvent flow rate of 1821.2 t/h was estimated. This flow rate was then used as an initial input in Aspen Plus to obtain a converged result in the next step, where the absorber was switched to the rate-based model. The column diameter (D) can be obtained either via Aspen Plus interactive sizing or estimated given the gas flow rate (G) and the superficial velocity of the gas (U_s) where more details can be found in Agbonghae et al. (2014) and Otitoju et al. (2023):

$$D = \sqrt{\frac{4G}{\pi U_s}} \qquad \text{Eq. (8)}$$

The diameter of the column was varied in Aspen Plus to ensure 80% flooding, where a column diameter of 7.8 m was obtained.

Step 3: Determining the cost optimal height of the absorber column – Sensitivity study on column height

To determine the cost optimal height of the absorber column, a sensitivity analysis was conducted by varying the column height while keeping the column diameter and solvent flow rate (from Step 2) constant. The objective was to minimize the total annual cost (TAC), which includes both the equipment cost of the absorber and the steam cost for regeneration. The TAC was then divided by the annual operating hours to obtain the average cost of operation per hour, ensuring consistency in the units. The minimum TAC was achieved at a column height of 37.8 m, as shown in Figure C1.

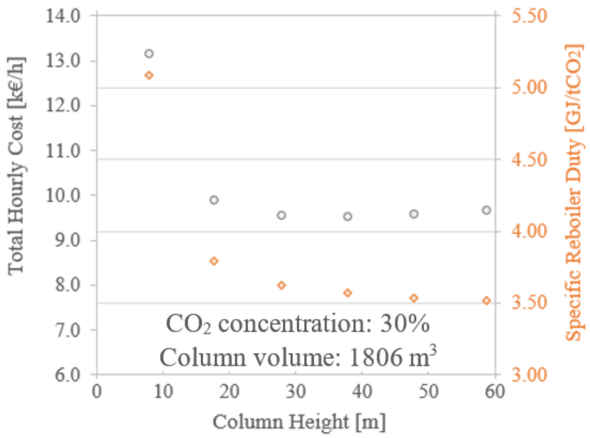

Figure C1. Effect of the column height on the specific reboiler duty and the total hourly cost at 30% CO_2 inlet concentration and at a fixed capture scale of 1Mt/y

Step 4: Determining the optimized solvent flow rate (lean loading) – with fixed column dimensions

With a fixed column diameter of 7.8 m and a height of 37.8 m (obtained at an initial solvent flow rate of 1821.2 t/h in Step 2), the solvent flow rate was then varied to identify the flow rate which optimize the specific reboiler duty. Figure C2 shows the result of the optimized specific reboiler duty of 3.542 GJ/tCO_2 obtained at a lean loading of 0.246 (corresponding to a lean solvent flow rate of 1945.5 t/h).

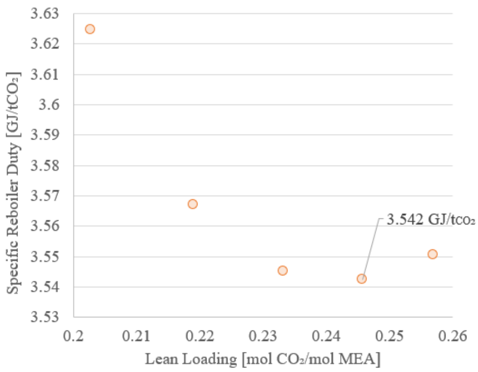

Figure C2. Lean loading Vs specific reboiler duty at 30% CO_2 inlet concentration and at a fixed capture scale of 1Mt/y

The solvent flow rate obtained from the sensitivity study differed by 6.8% from the initial estimate calculated with Eq. (7). As shown in Figures 8 and 9 in the paper, the reboiler duty and lean loading remain consistent across different capture scales when the CO_2 inlet concentration is fixed. Therefore, these results were applied to determine column dimensions at other capture scales with the same CO_2 inlet concentration. Details of the data points are provided in Table 13 and Appendix A (Tables A2–A4).

Appendix D: Correlation result details

Case study 1: Cement industry at 11.5% CO_2 inlet concentration (Table 16)

Table D1. Detailed CapEx and OpEx breakdowns for Case study 1

CapEx estimation method: DOE/NETL

MEA CC Process		**M€$_{2023}$**
Total Equipment cost (T.E.C)	Shortcut correlations	**42.07**
Supporting Facilities	71.4 % of TEC	30.04
Total labor cost	37% of (TEC + 71.4 % of TEC)	26.68
Bare Erected Cost (BEC)	**-**	**98.79**
Engineering& Construction management	10% of BEC	9.88
Engineering procurement construction (EPC)	**-**	**108.67**
Process contingency	25% of EPC	27.17
Project contingency	20% of (BEC + EPC + 25% of BEC)	46.43
Initial MEA cost	-	1.02
Total plant cost (TPC)	**-**	**183.29**
Owner's cost	15% of TPC	27.49
CC Process CapEx	**-**	**210.79**

Compression System (CS)		**M€$_{2023}$**
Total Equipment cost (T.E.C)	Eqn. (12) or Figure 11	**2.94**
Supporting Facilities	71.4 % of TEC	2.10
Total labor cost	37% of (TEC + 71.4 % of TEC)	1.86
Bare Erected Cost (BEC)	**-**	**6.90**
Engineering& Construction management	10% of BEC	0.69
Engineering procurement construction (EPC)	**-**	**7.59**
Process contingency	0% of EPC (Matured process)	0.00
Project contingency	15% of (BEC + EPC + 25% of BEC)	2.43
Total plant cost (TPC)	-	10.03
Owner's cost	15% of TPC	1.50
CS CapEx	**-**	**11.53**

Total CapEx	CC Process CapEx + CS CapEx	222.32
Total CapEx (Location factor applied)	Location factor in 2018: 1.1	**224.11**
Specific CapEx [M€$_{2023}$/t/h]	Operating hour of 8215 h/y	**2.63**

OpEx: MEA CC Process + Compression		**M€$_{2023}$**
Operating labour cost (OLC)	16 workers, Annual Salary: 50 000 $\$_{2018}$	0.94
Supervision	20% of OLC	0.19
Laboratory charges	15% of OLC	0.14
Maintenance and repair	5% of FCI*	2.98
Operating supplies	0.9% of FCI*	0.54
Local taxes insurance	2 % of FCI*	1.19
Plant overhead expenditure	(70.8% of OLC) + 3.6 % of FCI*	2.81
Administrative expenditures	17.7% of OLC + 0.9% of FCI*	0.70
Fixed OpEx	Sum of above	9.49
Fixed OpEx (Location factors applied)	Labour location factor in 2018: 1.82	**11.74**
Variable OpEx	Shortcut correlations	**44.39**
Total OpEx	Variable OpEx + Fixed OpEx	**56.13**

*The fixed capital investment (FCI) is calculated as described by Turton et al. (2018)

The location factors are updated from a base year (e.g. 2018) to a reference year of TEA study (e.g. 2023) as follows (Peters et al., 2018):

$$\text{Location Factor}_{2023} = \text{Location Factor}_{base\ year} \times \left(\frac{\text{Exchange Rate}_{2023}}{\text{Exchange Rate}_{base\ year}}\right) \quad \text{Eq. (D1)}$$

Case study 2: Cement industry at 22% CO_2 inlet concentration (Table 17)

Table D2. Detailed CapEx and OpEx breakdowns for Case study 2

CapEx estimation method: Roussanaly et al. (2018)

MEA CC Process		**M€$_{2023}$**
Total Equipment cost (T.E.C)	Shortcut correlations	**37.60**
Supporting Facilities	71.4 % of TEC	26.85
Total labor cost	37% of (TEC + 71.4 % of TEC)	23.85
Total Direct Cost (TDC)	**-**	**88.29**
Engineering& Construction management	14% of TDC	12.36
Engineering procurement construction (EPC)	**-**	**100.65**
Total plant cost (TPC)	19% of EPC + EPC	**119.77**
Initial MEA cost	-	2.44
Owner's cost	15% of TPC	17.97
CC Process CapEx	**-**	**140.19**

Compression System (CS)		**M€$_{2023}$**
Total Equipment cost (T.E.C)	Eqn. (12) or Figure 11	**3.06**
Supporting Facilities	71.4 % of TEC	2.19
Total labor cost	37% of (TEC + 71.4 % of TEC)	1.94
Bare Erected Cost (BEC)	**-**	**7.19**
Engineering& Construction management	10% of BEC	0.72
Engineering procurement construction (EPC)	**-**	**7.91**
Process contingency	0% of EPC (Matured process)	0
Project contingency	15% of (BEC + EPC + 25% of BEC)	3.38
Total plant cost (TPC)	-	11.29
Owner's cost	15% of TPC	1.69
CS CapEx	**-**	**12.98**

Total CapEx	CC Process CapEx + CS CapEx	153.17
Total CapEx (Location factor applied)	Location factor in 2018: 1.01	**141.77**
Specific CapEx [M€$_{2023}$/t/h]	Operating hour of 8400 h/y	**1.50**

OpEx: MEA CC Process + Compression		**M€$_{2023}$**
Operating labour cost (OLC)	Based on Nwaoha et al. (2018)	1.81
Supervision	20% of OLC	0.36
Laboratory charges	15% of OLC	0.27
Maintenance and repair	5% of FCI*	2.77
Operating supplies	0.9% of FCI*	0.50
Local taxes insurance	2 % of FCI*	1.11
Plant overhead expenditure	(70.8% of OLC) + 3.6 % of FCI*	3.27
Administrative expenditures	17.7% of OLC + 0.9% of FCI*	0.82
Fixed OpEx	Sum of above	10.90
Fixed OpEx (Location factors applied)	Labour location factor in 2018: 1.82	**14.34**
Variable OpEx	Shortcut correlations	**29.06**
Total OpEx	Variable OpEx + Fixed OpEx	**43.40**

*The fixed capital investment (FCI) is calculated as described by Turton et al. (2018)

Case study 3: Refinery at 8.2% CO_2 inlet concentration (Table 18)

Table D3. Detailed CapEx and OpEx breakdowns for Case study 3

CapEx estimation method: IEAGHG (2017)

MEA CC Process		**M€$_{2023}$**
Total Equipment cost (T.E.C)	Shortcut correlations	**25.34**
Supporting Facilities	58.8 % of TEC*	14.90
Bare Erected Cost (BEC)	**-**	**40.25**
Engineering& Construction management	20.1% of BEC*	8.11
Engineering procurement construction (EPC)	**-**	**48.35**
Project contingency	10% of EPC*	4.84
Total plant cost (TPC)	**-**	**53.19**
Owner's cost and others	25.9% of TPC*	13.79
CC Process CapEx	**-**	**66.98**

Compression System (CS)		**M€$_{2023}$**
Total Equipment cost (T.E.C)	Eqn. (12) and/or Figure 11	**1.60**
Supporting Facilities	58.8 % of TEC*	0.94
Bare Erected Cost (BEC)	**-**	**2.54**
Engineering& Construction management	20.1% of BEC*	0.52
Engineering procurement construction (EPC)	**-**	**3.06**
Project contingency	10% of EPC*	0.31
Total plant cost (TPC)	**-**	3.36
Owner's cost and others	25.9% of TPC*	0.87
CS CapEx	**-**	**4.24**

Retrofitting/ Interconnecting (RI) cost	Sannan et al. (2017)	91.65
Total CapEx	CC Process CapEx + CS CapEx + RI	162.87
Total CapEx (Location factor applied)	Location factor in 2018: 1.01	**157.57**
Specific CapEx [M€$_{2023}$/t/h]	Operating hour of 8003 h/y	**4.38**

*The percentages are obtained from the cost results obtained from Sannan et al. (2017)

OpEx: MEA CC Process + Compression + Retrofitting		**M€$_{2023}$**
Operating labour cost (OLC)	10 workers, 80 000 \$$_{2016}$ (Sannan et al., 2017)	1.05
Annual maintenance	2% of TPC/ 60%*	1.89
Others	0.5% of TPC	0.28
Fixed OpEx for Retrofitting	Sannan et al. (2017)	1.41
Fixed OpEx	Sum of above	4.63
Fixed OpEx (Location factor applied)	Location factor applied to CapEx-related items	**4.36**
Variable OpEx	Shortcut correlations	**16.49**
Total OpEx	Variable OpEx + Fixed OpEx	**21.44**

*60% is the share of annual material maintenance cost in the overall annual maintenance cost (Sannan et al., 2017)